

Piping and Instrumentation Diagram Development

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Moe Toghraei

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This book is dedicated to the soul of my father:

Behrouz

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Preface

The need for a book on piping and instrument diagram (P&ID) development is always felt in the process industries. However, for a long time there no book covered this topic.

There are several reasons for this.

One reason is the practical nature of this skill. A large number of books written for working professionals are authored by university professors. The skill of P&ID development however is not pure theoretical knowledge. It is a combination of technical skills and other considerations like ease of operation, ease of maintenance, client preferences, and jurisdictional codes.

On the other hand, the required technical skill is not exclusively in the territory of chemical engineers or instruments/control engineers or any other single engineering discipline. The set of skills for P&ID development comes from different engineering disciplines.

Because of the above issues, the P&ID development skill is always considered as an “on the job learning” skill.

The roadblocks of a book on P&ID development are not only the vastness of the skill or the practical nature of it, but also some preventing beliefs.

Some people claim that P&ID drawing is similar to a painting, which involves a bunch of creativity. Therefore “P&ID development” cannot be taught. However, the answer is in their question: even though painting needs a bunch of creativity it doesn’t prevent teachers/instructors from writing books to explain the fundamentals of painting and also showing some of the creativities by other painters to spur the learner’s creativity.

Some other people claim that P&ID development can be done only by following the company guidelines and such a topic cannot be taught as a general course. However, all the instructions in the company’s guidelines have underlying logic. This book tries to explain these logical backgrounds.

The goal of this book is to provide information about the development of P&IDs for designers and personnel of process plants.

When it comes to P&ID, there are three main group of knowledge may come to mind. They are:

- 1) The group of information on the technical “development” of P&IDs
- 2) The group of information that shows different elements on P&ID sheets
- 3) The group of information about how to draw already-developed P&IDs (drafting P&IDs).

A P&ID should first be developed (step 1), then drafted (step 3) based on the rules of P&ID appearance (step2).

This concept is shown in the diagram.



This book doesn’t address how to “draw” a P&ID (group 3). There is plenty of software and plenty of training courses by the software companies that cover that topic.

This book mainly focuses on the development of P&IDs and how to show different elements on P&IDs.

The information in this book will not only help in the development of P&IDs, but will also help in understanding the activities of process plants that are related to P&IDs.

Chemical engineers will use this book to learn how to design process plants based on selected and designed/specified equipment or unit operations or process units. They will learn how to “tie” together different units to make sure the plant runs safely and produces the predetermined products with the highest level of operability.

Chemical engineers and other engineers in process plants will use this book so that they can read and interpret P&IDs deeply in order to maintain any piece of equipment in the plant and/or doing repair.

There are several disciplines involved, including chemical engineering, mechanical, piping instrumentation and control, electrical engineering, and civil engineering disciplines.

All disciplines involving a process plant should be familiar with P&IDs.

Since the P&ID is a multidisciplinary drawing, the concepts must be presented in layman's terms in order to be accessible to a wide range of engineers. As it has been seen that individuals with different level of study, from engineers to technologists and technicians, have the duty of P&ID development, this book is written for whoever has enough knowledge of process plants and wants to learn P&ID development skills. Therefore the concepts are not necessarily explained in university level language.

The skeleton of this book has five parts as below.

Part 1: Fundamentals of P&ID Development

This part covers the fundamentals of P&IDs and P&ID development. Chapters 1 to 5 comprise Part 1 of this book.

At the beginning I will explain the nature and importance of P&IDs (Chapter 1). Then I will explain the milestones in developing P&IDs (Chapter 2). In Chapter 3 the “court of game”, or different sections of a P&ID sheet will be explained. In Chapter 4, the basic rules of drafting P&IDs will be discussed. Chapter 5 talks about the thought process for developing P&IDs and what goals a designer needs to look for to develop a good P&ID.

When talking about “piping and instrumentation diagrams” it seems the topic can be explained by explaining two elements of piping (and equipment) and instrumentation. However for different reasons I have decided to divide the topic into the three elements of pipes and equipment, instrumentation/control systems, and utility generation and networks.

Part 2 is devoted to pipes and equipment, Part 3 will cover instrumentation and control systems, and Part 4 covers topics related to utilities.

For each of these elements the skills for P&ID development is explained together with plenty of general practices for each component.

Part 2: Pipes and Equipment

The majority of process items (pipes and equipment) in different P&IDs are pipes and pipe appurtenances, valves (manual and automatic), containers (tanks and vessels), fluid movers (including pump, compressor, fan, and blower), and heat exchangers.

Part 2 has seven chapters. In Chapter 6, pipe and pipe fittings are discussed. Chapter 7 belongs to different types of valves.

Chapter 8 provides information about the development of P&IDs considering inspection and maintenance. As such provision needs to be made for specific types of pipe and valve arrangement and this topic is placed after Chapter 6.

Chapter 9 discuss different types of containers including tanks and vessels and the way we develop their P&IDs.

Chapter 10 covers fluid movers. Fluid movers include liquid movers or pumps and gas/vapor movers or compressors, blowers, and fans.

Chapter 11 talks about heat transfer units. They are mainly divided into heat exchangers and furnaces (fired heaters).

Pressure safety devices (PSDs) are discussed in Chapter 12. Although one main portion of PSDs are pressure safety valves (PSVs), and are a special type of valves, it was decided to devote a separate chapter to them. The reason is that another portion of PSD is rupture disks, which are not a type of valve, and also the concept of PSDs is adequately important to consider a separate chapter for them.

Part 3: Instrumentation and Control

Part 3 comprises the four Chapters of 13, 14, 15, and 16.

Chapter 13 developed to give a basic practical idea about instrumentation and control to the reader. As is mentioned there, the control system, or in a more complete phrase integrated control and safety, in each plant has three main elements.

In Chapter 14 the concept of control loop and the method of developing control loops on P&ID are discussed.

The first element of control is covered in Chapter 14 as “plant control”. The other two elements, interlock and alarm systems are covered in Chapters 15 and 16.

Part 4: Utilities

In Chapter 17, the reader will learn about utility systems in a process plant and how to develop their P&IDs. When talking about utilities, there are two separate concepts that should be discussed: utility generation and then the distribution of utilities and the collection of “used” utilities. Both of them are discussed in this chapter.

Part 5: Additional Information and Wrap-up

Part 5 covers additional information to that covered in the previous chapters. Part 5 has two chapters, Chapters 17 and 18.

Chapter 17 covers some additional small systems (tracing and insulation, utility stations, safety showers and eye washers, sampling systems, and corrosion coupons) and also an important topic that is very important in P&ID development.

The important topic, covered as part of chapter 17, is “design pressure and temperature considerations”. This topic covers precautions should be taken when tying together different process elements in P&IDs.

In chapter 18 some units that could be categorized in the previous chapters are presented. The important concept of design temperature and design pressure is also studied here.

Chapter 19 could be considered as summary of the previous chapters. In this chapter a general methodology is provided for P&ID development of a new item (not familiar for the designer) and then P&ID development of some common systems (like chemical injection system,

silo and solid transfer, etc.) is brought. At the end P&ID reviewing and checking is discussed.

Introduction: What is P&ID Development Skill?

The first thing is to decide is the meaning of P&ID development. The answer can be prepared regarding two aspects: the depth and the breadth.

All the items on a P&ID sheet went through two steps and several engineering disciplines. The depth of P&ID development could be defined as all activities to develop a P&ID but beyond the design.

The breadth of P&ID development could be defined as all activities by different engineering disciplines.

The depth of P&ID development is explained in more detail below.

Each discipline does the design and then P&ID development.

The design, in this context, means sizing or specifying a piece of an element such as a pipe, equipment, instrument, etc.

There are different disciplines involved in the design of a process plant, including process engineering, piping and piping engineering, instrumentation and control engineering, mechanical engineering, and civil engineering.

The duties of the chemical engineer in a CPI project can be broadly split into two categories: equipment sizing/specification and P&ID development. Therefore the chemical engineer needs to have skills in both classes.

The former skill needs primarily knowledge of hydraulic calculations, pump/compressor sizing, vessel/tank sizing, PSV sizing, and heat exchanger sizing. At a higher level of sizing skill the chemical engineer should have the knowledge of designing different unit operation and unit processes. For example an engineer in the air purification industry may need to know about the design of different solid–gas separation units. Another engineer in the oil refining industry may need to know how to size a distillation tower. All of these sizing skills could have been learned during the acquisition of an engineering degree.

However, the latter—which is not formally taught and considered as “on the job” learning—includes the skills required to determine appropriate piping, piping appurtenances, proper tanks, vessels, pumps, and also instrumentation/control to determine the goal of the plant.

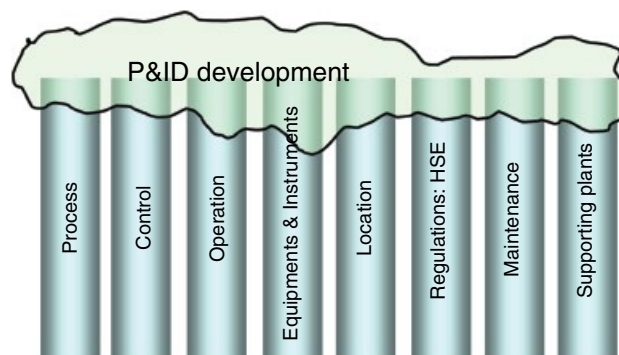
Equipment sizing is beyond the scope of this book. However, sometimes it is not easy to put a separation line between equipment sizing and P&ID development. The result of equipment sizing AND P&ID development effort makes P&IDs.

Some concepts in this book may be considered as sizing skill concepts by some individuals but they are still included in the book. The reason is that a P&ID document could be used to “check” the sizing during the P&ID development stage. As a P&ID gives a big picture of a plant, sometimes the mistakes in equipment sizing could be revealed when the equipment appears on the P&ID. Therefore, some of the concepts in this book can be used to “roughly” check the accuracy of sizing of pipes and equipment.

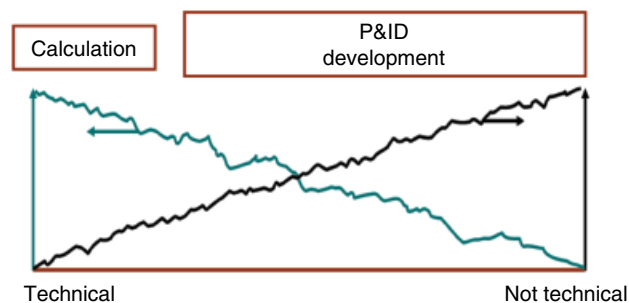
The breadth of P&ID development is explained in more detail below.

The question is: which sector of P&ID skills are outlined here? Chemical engineering sector? Instrumentation and control sector? Mechanical engineering sector?

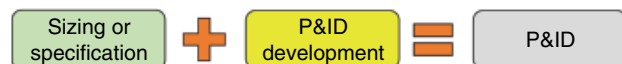
As was mentioned before, P&ID development skills are stretched over multiple disciplines. A P&ID sheet is the result of exhaustive work by different disciplines. Each item on a P&ID may be rooted in a deep concept of chemical engineering, instrumentation and control engineering, mechanical engineering, etc.



At the beginning of P&ID development the items are added on P&IDs based on some calculations and sizing. However, when we progress further, the added items on P&IDs are not backed by quantitative documents but by qualitative, judgment-type decisions.



The content of this book covers not only the chemical engineering sector of P&ID development but also the required knowledge of other disciplines to develop P&IDs is covered.



This book, however, doesn't eliminate the need for professionals in different areas because it only provides some rule of thumbs in those areas to help and accelerate the developing of P&IDs.

No attempt was made to explain the deep concepts, as they are discussed in other books with better depth and

breadth. This is the reason that there are few references in each chapter because I try to convey only the skill of P&ID development.

Therefore, this book can be considered as book of rules of thumb for P&ID development. The readers learn the "root" of knowledge needed to refer to the respective resources.

Acknowledgement

In preparing this book I indebted to many professionals who have helped me. Amongst them, I acknowledge assistance of Dr. Ashgar Mesbah for his helpful notes on Chapter 13, Greg Pajak for his consultation to provide a better content in chapter 16. I am also very grateful my old friend, Shahriar Gilanmorad for accepting to develop and draw many schematics of this book.

I also must acknowledge all the helpful guidance and help I received from Rasoul Sayedin during my career and regarding P&ID development.

Finally, to my wife, Mishga Abedin, who suffered but supported me through the writing of this book.

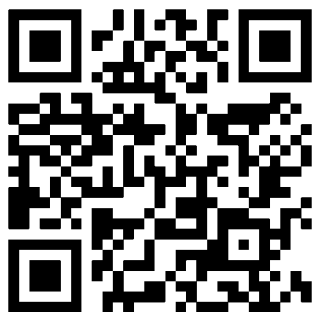
At the end, I will appreciate it very much if all professionals who use this book let me know of any errors which is made or any change, they believe needed to improve the book.

Moe Toghraei
Calgary, Alberta
2018

About the Companion Website

This book is accompanied by a companion website:

www.wiley.com/go/Toghraei_PID



The website includes:

- Figures for which user should mention their comments or an interpretation.

Part I

Fundamentals of P&ID Development

In part 1 we are going to cover the common rules of P&ID development. This part has five chapters:

Chapter 1: What is P&ID?

Chapter 2: Management of P&ID development

Chapter 3: Anatomy of a P&ID sheet

Chapter 4: General rules in drawing P&IDs

Chapter 5: Principles of P&ID development

In Chapter 1 we will cover the identification of P&ID and its role in process industries.

Chapter 2 covers the progress steps of P&ID during a design project.

Chapter 3 talk about different components of a P&ID sheet and their meaning.

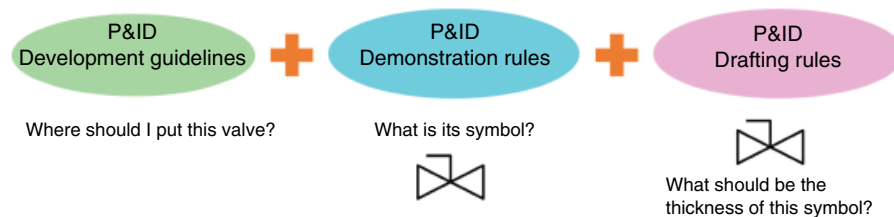
Chapters 4 and 5 cover some rules and guidelines about P&IDs. In the world of P&IDs there would be at least three sets of rules, guidelines or standards.



An example is a valve. The engineer should decide if they need to put the valve in that specific location or not, and if needed, which type of valve with which type of actuator should be used.

The engineer uses “P&ID development rules.”

In the second step, when it is decided to put a manual gate valve, the question is how to show it. The engineer and the drafter together agree on a specific symbol based on “P&ID demonstration rules.”



As it was mentioned before, we are not going to talk about drafting rules. Chapter 4 covers demonstration

rules and Chapter 5 explains the general guidelines about P&ID development.

1

What Is P&ID

1.1 Why Is P&ID Important?

The piping and instrumentation diagram (P&ID) is what might be considered the bible of a chemical process plant (CPI). It provides a lot of information for the manufacturing of the equipment, installation, commissioning, start-up, and the operation of a plant. It also presents how a process plant should handle emergency situations.

The P&ID is a frequently referenced document throughout a project term – from the designing stages to the plant-in-operation phase – by various engineering disciplines, in technical meetings with vendors or manufacturers, hazard and operability study (HAZOP) meetings, management meetings, and project scheduling and planning. It is also one of the few documents created by multiple engineering groups such as Process, Instrumentation and Control (I&C), Plot plant and Piping (PL&P), Mechanical, Heat Ventilation and Air Conditioning (HVAC), and to a lesser extent Civil, Structural, and Architecture (CSA) and environmental or regulatory groups.

The information provided by the P&ID allows for the generation of other documents, including piping isometric drawings, the piping model, equipment and instrument lists, cause-and-effect diagrams, control philosophy, alarm set-point tables, line designation table (LDT) or line list, material take-offs, loop diagrams, tie-in lists, and many more (Figure 1.1). The P&ID can ironically be considered an acronym for “primary interdisciplinary document.”

All the abovementioned groups involved in a process plant should be familiar with P&IDs to some extent. P&IDs can be considered the source document to prepare “shopping lists” for piping, mechanical, and I&C requirements.

Process and I&C groups should be fully knowledgeable about P&IDs because they are the main developers of P&IDs, together with Mechanical and Electrical groups who can also provide some input during development such as the equipment list from mechanical engineers

and motor list and electrical classification list from electrical engineers. I&C practitioners should be familiar with the P&IDs because from it they develop their instrument list, I/O list (a document listing instrumentation which serves as an input or output of the control system), and so on. The Piping group should be familiar with P&ID because it is their main working document and they need this to develop their piping model and piping isometric drawings. Civil engineers should also know P&IDs, although to a lesser extent. However, if the materials are concrete, like in large wastewater treatment plants, they should be familiar with P&IDs.

In HAZOP meetings, the discussion is mainly on P&IDs and about different potential hazards in a designed plant and the required safeguards.

In value engineering meetings, again P&IDs have an important role. Different elements of a designed plant are discussed from cost-savings point of view, and if one item is not really necessary, it will be removed. A deep understanding of P&IDs is needed, which can be provided by this book.

One important factor when designing a process plant is the quality of its P&ID. The more complete a P&ID is, the more developed the project is. P&IDs are also used in project schedule meetings.

Therefore different groups can add information to a P&ID and also use it for their activities (Figure 1.2).

During construction and installation of process plants, P&IDs are important, too. Although the majority of construction activities are done based on documents other than P&IDs, having the most updated P&ID on the site helps the constructors eliminate any vagueness when questions arise. P&IDs have been used during construction to solve a wide variety of problems, from tank nozzles to insulation and painting. However, where well-prepared documents are available, there is less need for the P&ID during construction.

During the operation of process plants, the main drawing referenced is a P&ID. P&IDs should be always in the plant and updated based on the latest changes.

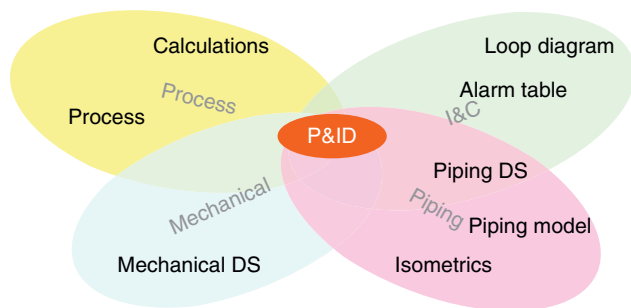


Figure 1.1 The P&ID is used by other groups to prepare other project documents.

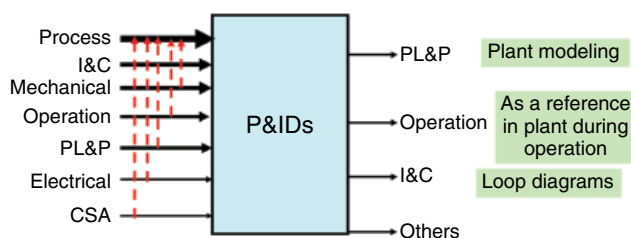


Figure 1.2 The P&ID is a document consolidated and used by different groups.

P&IDs are used by operations personnel, control technicians and engineers, maintenance personnel, and other stakeholders. One main use of P&IDs is for maintenance personnel to initiate lockout–tag out actions. This concept will be discussed in Chapter 8.

Some individuals in the operation of a process plant may consider to not know about the development of P&IDs because it is “not their business.” However, this approach is not completely correct for different reasons. For example, a considerable number of items on P&IDs are things inherited from the design and development stages of the P&ID; therefore, to have a good understanding of the P&ID, its development needs to be understood.

Sometimes an “abridged” version of a P&ID is created for the purposes of operation. Some people try to create an “Operational P&ID’s” because they claim that the P&IDs they receive (by the time the plant is in operation) has many features related to the design phase of project that are not relevant to the plant’s operation. The concept of an Operations P&ID is not accepted by all industry professionals.

1.2 What Is a P&ID?

A P&ID is the focal drawing in all process plants. P&IDs may be named differently by each company; however, P&ID is the most common. P&IDs can also be called

engineering flow drawing (EFD) or mechanical flow diagram (MFD).

A process plant can be an oil refinery, a gas processing plant, a food processing plant, mineral-processing plant, pulp-and-paper plant, pharmaceutical or petrochemical complexes, or water and wastewater treatment plants.

All the plants that make non-discrete “products” use P&IDs to show their process. For example, in an automotive factory, they make discrete things (e.g. cars), so they do not use P&IDs.

Some other industries that traditionally are not classified as process industries have started to develop and use P&IDs. One such example is the HVAC industry.

P&IDs can even be used to show the system of some machines that do some processing of some sort.

Do you mean that can I draw a P&ID for my washing machine, vacuum cleaner, or even coffee maker? Yes, you can, and I did it as practice. However, it is not helpful during the design stage of the “project” or for household repair specialists.

P&ID is a type of engineering drawing that describes all the process steps of a process plant. It basically is a process plant on a paper. A P&ID is a schematic diagram of pipes, process equipment, and control systems by a set of predefined symbols with no scale and no geographical orientation. Equipment symbols are typically a side view of the real shape of the equipment, and if possible, are shown relative to their actual sizes.

Different types of lines on the P&ID represent pipes and signals. However, the length of lines do *not* represent the real length of pipes or signal carriers (e.g. wires).

There are, however, a set of P&IDs that are shown in plan view rather than in side view. They are generally drawings that only show piping. Drawings, such as utility distribution P&IDs, are shown as schematics but in plan view (Figure 1.3). Different types of P&IDs will be discussed in Chapter 4.

1.3 P&ID Media

P&ID is handled in two different platforms: paper media and electronic media. P&IDs used to be outlined on paper. We are now in a transition state and moving to electronic P&IDs. Whether there will be paper P&IDs will still needs to be determined.

P&IDs are published on paper. The paper size is different for each company because P&IDs are not drawn to scale. The only criteria in choosing a paper size for a P&ID are the ease of reading its content and the ease of handling.

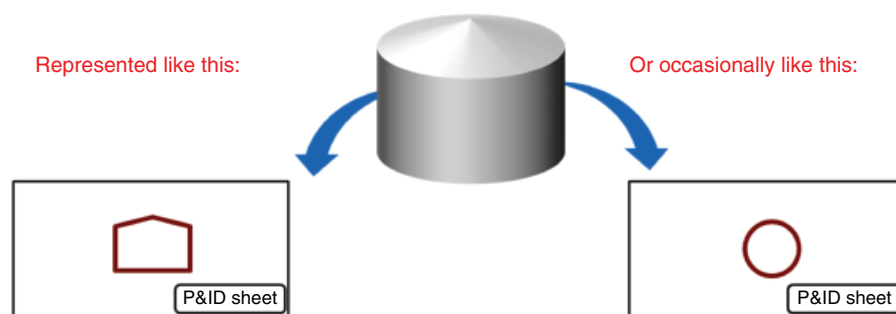


Figure 1.3 Each element is represented through a symbol on the P&ID.

I remember there were days that we used P&IDs in big roll sheets. We did not miss any streamline, but it was difficult to find a suitable machine to reproduce them!

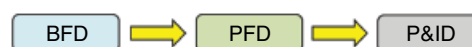


Figure 1.4 Root of P&ID.

Some companies have defined a “standard” P&ID sheet size to specify the size of P&ID sheets that will be handed over to the client. This sheet size could be “D” size (in American system) or “A1” in ISO system.

Official markups should go on the official P&ID set.

Designers may use a smaller P&ID sheet size (e.g. “B” or “A3” size) as a personal copy if they find it difficult to work with the actual size.

Electronic P&IDs or e-P&IDs are gaining more popularity not only because of their ease of transfer but also because of added “smart” capabilities. Smart electronic P&IDs can be used by operators on their tablet (approved by plant management), and the operators can easily find different items on the e-P&IDs. e-P&IDs are also beneficial during the design and development stages. In good P&ID development software, there are tools to capture comments and markup by different people. The parts of a P&ID can be snapshots and used in management of change documents.

1.4 P&ID Development Activity

A P&ID is developed based on the information from process flow diagram (PFD), which is developed based on a block flow diagram (BFD). This is depicted in Figure 1.4.

The BFD is the preliminary document in the development of a project and outlines the basics and general information of the project. The PFD is actually the expanded view of the BFD. It is the job of the designer to add further details to the PFD design before the final document – the P&ID – is developed. The BFD and PFD only show the main elements of the plant, whereas the P&ID shows more detailed elements. While BFD and PFD are normally considered internal documents of process engineers, P&ID is a cross-discipline document.

The development of BFD and PFD requires exhaustive studies and rigorous calculations or simulations.

Going through these preliminary drawings is a must because each decision for main items has an impact on the project. However, P&ID development does not merely moving from PFD to P&ID. A PFD only covers the main items of a plant. There are several other items in a plant that do not appear on a PFD (e.g. sampling systems and HVAC systems). For these items, P&ID development basically means developing the BFD, the PFD, and then the P&ID. Although the BFD and the PFD of these items are not always drawn, designers should, at least, visualize them in their mind before trying to develop the P&ID.

Therefore, P&ID development is the activity of evolving the P&ID from the PFD for main items *and* for non-main elements going directly on a P&ID at the outset.

Because P&ID development starts from the PFD, the symbols and text information on the PFD should be transferred to the P&ID, but this transformation is not a blind act. The symbols and text information need to be converted to their corresponding symbols and text data on the P&ID.

Let’s start with symbols and then we will talk about text information.


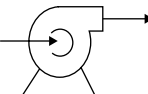
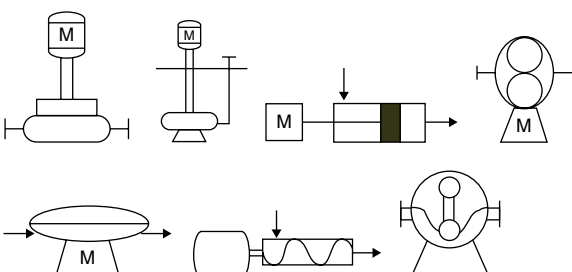

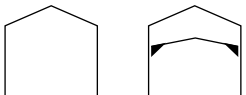
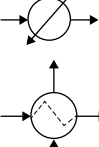
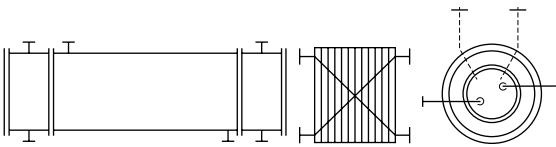

It is important to point out that the mechanical details of a plant are *not* shown on a PFD, only the process steps or process elements of the plant are shown. For example, even though we see a  on a PFD, we do not introduce a tank as hardware here. What we are trying to say at the PFD level is, “we have something here that stores the liquid of interest.” It could be a fixed roof tank or floating roof tank, but in this stage of the project we do not know yet or we do not care to show it on the PFD. Because of this, the physical dimensions of the tank on a PFD are not mentioned; we only say that the normal storage liquid capacity of this tank is, for example, 500 cubic meters.

Table 1.1 PFD symbols compared with P&ID symbols.

PFD symbol	Symbol
	
	
	

In a PFD, we see a  only as one process step, although such a symbol brings to mind the concept of a tank.

Also in some PFDs, there is only one symbol for different types of pumps including centrifugal or PD type (and this could be confusing for those who are not familiar with a PFD). When we show a symbol of a pump on a PFD, we are basically saying, “the liquid must be transferred from point A to point B, by an unknown type of pump that will be clarified later in the P&ID.” A pump symbol in a PFD only shows “something to transfer liquid from point A to point B.” Each symbol on a PFD has a general meaning and does not refer to any specific type of that equipment (Table 1.1).

The other aspect of the concept of a PFD is that its lines do not necessarily represent pipes; they only represent streams. This means that one line on a PFD could actually represent two (or more) pipes that go to two (or more) parallel units.

Therefore, one main activity during P&ID development is deciding on the type of a piece of equipment. A general symbol of a heat exchanger on a PFD should be replaced with a specific symbol of shell and tube heat exchanger, a plate and frame, or the other types of heat exchangers when transferred to the P&ID. Sometimes the type of process elements is decided when sizing is done, but in other times this decision is left to P&ID developers.

This activity is commonly overlooked because in many cases P&ID development starts with a set of go-by P&IDs rather than blank sheets.

Regarding text information, the main difference is in equipment callouts, which will be explained in detail in Chapter 4. Here, however it can be said that an equipment callout is a “box of data” about a piece of equipment shown on a PFD and on a P&ID.

In a nutshell, the difference between equipment callouts on a PFD and on a P&ID is that on the PFD the data are operational information, while on the P&ID, the data are mechanical information. As a PFD is mainly a process engineering document, all its information are for normal operations of a plant. The P&ID is a document that illustrates the capability of the equipment. Table 1.2 shows the differences between one-item callouts on a PFD and on a P&ID.

As can be seen in Table 1.2, the numerical parameters in a PFD callout are generally smaller than the corresponding parameters in a P&ID callout because on a PFD, only the “normal” value of a parameter is mentioned, whereas in a P&ID the “design” value of the parameter is reported.

There are, however, debates on the parameters that should be shown on a PFD callout. From a technical point of view, a PFD callout should not have the sparing philosophy, and also the reported capacity should be the capacity of *all* simultaneously operating parallel units. But not all companies agree on this.

Table 1.3 summarizes the differences between PFDs and P&IDs.

Table 1.2 A PFD callout compared with a P&ID equipment callout.

Equipment	Typical PFD equipment callout	Typical P&ID equipment callout
Tank	300-T-120 Wash water tank Capacity: 735 m ³	300-T-120 Wash water tank Size: 9850 mm DIA × 9000 mm H Total capacity: 865 m ³ Design: 1.25 kPa(g) @ 90 °C 0.25 kPa(g) VAC @ 90 °C Material: C.S. MDMT: −45 °C Insulation: 38H Trim: TT-419054-DAB
Pump	P – 123A/B Wash water pump Capacity: 664 gpm Sparing: 2 × 50%	P – 123A/B Wash water pump Rated spec: 730 gpm @ 206 psid Drive power: 300 hp. Material casing / impeller: CS / 8Cr Trim: A-342 Sparing: 2 × 50%

Table 1.3 Comparing the differing traits of the PFD and the P&ID.

		PFD	P&ID
Goal during design project		Process engineering internal document, for management decision, to get regulatory licenses, etc.	Link between Process group and other design groups
Goal during plant operation		To give a general idea of the plant to the plant operation team including engineers	Detail of plant for Operation and Maintenance teams
Concept		Showing process steps and some control loops (not interlocks)	Showing mechanical items and all control loops and interlock functions
“Look”		Different symbols that are connected to each other by some lines that represent “streams” (Your gut feeling may tell you what the symbols represent, but you need to refer to the PFD legend sheet to be sure.)	Generally a more crowded diagram Different symbols that are connected to each other by some lines representing “pipes” (Your gut feeling may tell you what the symbols represent, but you need to refer to the P&ID legend sheet.)
Content	Equipment	<ul style="list-style-type: none">● Only a general idea of each piece of equipment● No attempt is made to show mechanical detail for the equipment in the symbol and in callout information● The intent is to show the information that is mainly related to the heat and material balance (H&MB; i.e. volumes, flow rates, heat duty, and not the head of pumps or heat transfer areas of heat exchangers.)● Equipment is tagged	<ul style="list-style-type: none">● All process equipment is shown with maximum detail● Mechanical details that are not directly related to process may not be shown (e.g. gearboxes)● Electrical details that are not directly related to process may not be shown, but electric motors sometimes are shown
	Pipes	<ul style="list-style-type: none">● There are no pipe per se; “lines” represent streams and <i>not</i> pipes● Streams are tagged	All pipes are shown in the form of lines and are tagged
	Instruments	Instruments are <i>not</i> tagged	All instruments are shown and are tagged
	Control	Major control loops are shown	Complete control loops are shown and all of their elements are tagged
	Interlock	Only a few shutdown valves may be shown	All safety-instrumented functions are shown completely or briefly
	Alarm	No alarms are shown	Alarms symbols are shown

2

Management of P&ID Development

2.1 Project of Developing P&IDs

If the development of the P&IDs is viewed as a project (inside of the main plant design project), it has some features that need to be addressed to successfully complete it. The questions that should be answered are: Who should develop P&ID? How many man-hours is needed? Which quality should be followed?

These questions are answered in this chapter.

2.2 P&ID Milestones

P&ID development is a smooth evolution of a P&ID to the completion revision. However, because of different reasons, this smooth movement is split into different steps by project gates. These project gates are checkpoints to evaluate the set of P&IDs, to make sure that the project goes toward the direction it was intended, and to check the cost of the project for go or no-go decision.

A P&ID sheet may start as blank, but more commonly, it starts with a “go-by sterilized” diagram. A go-by sterilized diagram is a P&ID with some elements of the design from a previous similar project. An engineering company with experience in a specific type of design may decide to start the P&ID development activities from a similar project P&ID after removing the confidential aspects and details of the drawing. The P&ID is stripped of confidential information and is known as a “sterilized” diagram.

What is shown on the P&ID in the early stages of a project is a vague idea of a plant, whereas in the later stages of the project, the P&ID can finally be used for the construction of the plant. This last revision of a P&ID in a design project is Issued for Construction, or IFC. The P&ID goes through different milestones during a project: Issued for Review (IFR), Issued for Approval (IFA), Issued for Design (IFD), and IFC (Figure 2.1).

The first step is IFR. In this step, an engineering company has completed the primary P&IDs and lets the client

review them. These P&IDs represent the first thoughts about a plant and will be used as a starting point.

Some engineering companies issue one other revision of P&IDs before this milestone, and it is known as Issued for Internal Review, or IFIR. The objective of this IFIR is to have a set of P&IDs for review by the engineering company without the participation of the client. Not all engineering companies issue IFIR P&IDs even though they do need to internally review the drawings before allowing the client to look them over. Companies do this without officially issuing a P&ID as IFIR; they simply perform the internal review based on the latest copy of the P&IDs.

The next milestone of P&ID development is IFA. After the client has reviewed the IFR version of the P&IDs, they mark up the P&IDs with their thoughts and requirements and then return the marked-up P&IDs to the engineering company. The engineering company will then implement and address the client’s markups on the P&IDs; however, they do not do so blindly. For each individual markup, there may be many discussions, calculations, and studies to support the client’s markup or alternatives. After all of these activities are complete and the client’s markups have been implemented on the P&IDs (or some other agreed-upon solution has been established), the next revision of the P&IDs is ready to be issued; this is the IFA. At the end of this phase, the P&IDs are presented to the client, and the engineering company waits for client approval.

Clients may need some time for approval, and they may bring up new concerns that need to be addressed. Therefore, an IFR revision of P&IDs with the client’s markups will also then be returned to the engineering company. At this point, the engineering company starts to implement the client’s new markups (or convince the client on an alternative), and they also start adding more details to the P&IDs. At the end of this period, the P&IDs will be issued as IFD. The phrase *issued for design* seems weird. A question may arise: “Then what was the nature of all the activities before the IFD version of P&IDs?



Figure 2.1 The P&ID milestones.

Were they *not* design work?” However, the word *design* in IFD has a specific meaning. It means “design by groups other than Process discipline.” After issuing IFD P&IDs, the Process group lets other groups know that “my design is almost done and is firm, so all other groups can start their designs based on these (fairly) firm P&IDs.”

This is an important step because groups other than Process, including Instrumentation and Control, Piping, Mechanical, Electrical, and Civil can only start their (main) design based on a firm process design. If others start their design before a firmed-up process design, it may end up being costly because every change in the process design will impact other groups’ designs. However, it should be noted that after the issue of IFD P&IDs, it is not the case that process design is finished because process still continues its work but at a different and slower pace.

After IFD P&IDs, other groups do not expect the Process group to make big changes to the P&IDs.

All the steps up to the IFD version of P&IDs fall under basic engineering or front-end engineering and design (FEED), and all activities after IFD fall under detailed engineering.

One important activity that should usually be done before the IFD version of P&IDs is the hazard and operability study, or HAZOP. The HAZOP is an activity that seeks to identify flaws in design. It is a structured and systematic investigation technique to discover flaws in a specific process design. Generally, a HAZOP study is conducted in the form of a multiple-day meeting with people from different groups present.

The HAZOP study does not necessarily propose *solutions* to mitigate a process flaw; rather, it identifies the flaws and lists them in a HAZOP recommendation list. It is then the responsibility of the designer to address these flaws after the HAZOP meetings and close out the HAZOP issues.

In an ideal world, the HAZOP would be done before the IFD version of P&IDs because the HAZOP meeting may impact the process design heavily, and it is a good idea to keep all the big process changes handled before the IFD version of P&IDs. However, some companies decide to have HAZOP meetings after IFD P&IDs for different reasons, including a tight schedule or a lack of detailed P&IDs from vendors.

When a company wants to start a HAZOP study on a P&ID set, they may decide to do it on the latest and greatest version of the P&IDs, either officially issued or not. If the decision is to do the HAZOP on officially

issued P&IDs, the revision of the P&IDs is Issued for HAZOP, or IFH. Not all companies issue an IFH version of P&IDs for the purpose of the HAZOP study, and instead they do the HAZOP on the latest available P&IDs.

As was mentioned, all the activities after the IFD version of P&ID are part of detailed engineering. The client decides which activities should be done during the FEED stage of the project and which activities can be left for the detailed-engineering stage. A client can decide how complete a P&ID should be at each milestone. However, there is one thing that is almost universally accepted: There should be *no* contact with vendors during the FEED stage of a project, and all vendor contacts can start during the detailed-engineering stage to eliminate vendors’ involvement in process selection and design.

This also means that all the information on the P&IDs up to the IFD version comes from the engineering company’s experience and knowledge, and if there is a need for vendor information, the engineering company uses general vendor information or catalog information.

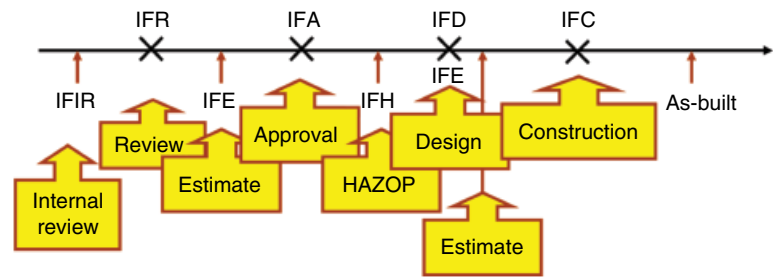
Later, during detailed engineering, all the assumed vendor-related information will be evaluated against the *actual* information provided by the selected vendor, and the information will be fine-tuned.

This concept shows the importance of previous experience for P&ID development.

There is one big exception to this rule and that is items with a long lead time. Long-lead items are the equipment whose delivery to site is long (maybe 2 years or more). For long-lead items, contact with the vendor can be started even during the early stages of the project or the IFR version of the P&IDs, which minimizes the impact of long-lead items on the project schedule. Long-lead items are generally the main equipment of a plant and are large or expensive ones. These may be different from plant to plant, but in general, equipment such as boilers, distillation towers, and furnaces can be considered long-lead items.

The next, and possibly last, P&ID milestone is IFC. Basically, from a P&ID point of view, the detailed-engineering activity consists of improving the P&ID from the quality of IFD to the quality of IFC.

As it was mentioned previously, during P&ID development, there could be several economic go or no-go gates put in place by the client. At each of these “gates,” the client needs a cost estimation report for the project to check if they want to continue the project, cancel it, or put it on hold. Therefore, there are usually three cost estimates during P&ID development. Each cost estimate can be done based on a copy of the P&ID set, or the client may ask for an official issue of the P&IDs for the purposes of cost estimation. For cost estimation purposes, an engineering company may issue P&IDs as Issued for Estimate, or IFE (Figure 2.2).

Figure 2.2 A more complete list of P&ID milestones.

In an ideal world, P&IDs will progress from IFR to IFA and finally IFC, but in reality there could be multiple interim milestones such as re-IFA, re-IFD, and multiple re-IFCs.

After building a process plant, what exists does not necessarily comply completely with the IFC version of the P&ID. There could be some constraints during the construction phase that force the designers to change some aspects of design and IFC P&ID. Therefore, a final P&ID should be prepared and issued based on the existing process plant. This revision of P&ID is called the “as-built revision.” An as-built P&ID could be prepared by collecting all the markups during the construction phase and implementing them in IFC version of the P&ID. If there are complicated changes in an area or there is a lack of thrust to markups on the IFC P&ID, the as-built P&ID can be prepared by a group of individuals who trace the different pipes and draft them. With the outsourcing tasks, an as-built drawing can be prepared remotely with the help of 360-degree cameras placed in suitable locations in a plant.

Some companies prefer not to use the phrase *as-built P&ID* and instead use the phrase of *as-recorded P&ID*. This preference is only due to legal concerns. By naming a set of P&ID *as-recorded* if an item is missed during observing or drafting, the engineering company will not be held liable.

It is important to keep the P&ID of an operating plant updated *always*.

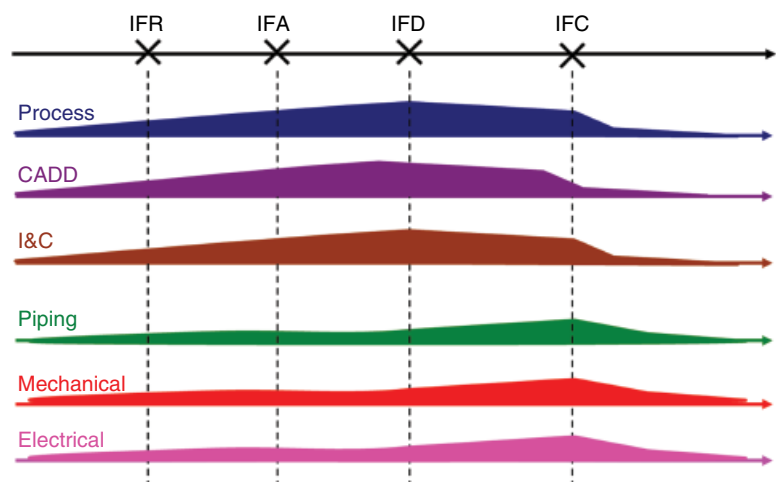
After commissioning of a process plant, there can be reasons to change things in the design such as troubleshooting, upgrading, or other projects. For such cases, a P&ID should be revised to reflect the new design. The designer normally starts with an as-built P&ID and then adds, deletes, or changes things in the P&ID.

2.3 Involved Parties in P&ID Development

As was discussed previously, the development of P&IDs is the result of team activities. People from different groups work together to develop a good quality P&ID set. However, their workload is not necessarily the same during a project (Figure 2.3). Generally at the beginning of a project, the Process group is heavily involved in the P&ID development, and later the other groups join in and help add to the P&ID. From the other side, the Process group is generally the first group that finishes their task and departs the development team.

The Instrumentation and Control (I&C) group works at almost the same pace as the Process group. However, I&C adds instruments and control systems on the items that the Process group listed on the P&IDs.

The computer-aided design and drafting (CADD) group has a load similar to the Process group, but it is a

Figure 2.3 Workload of different groups during P&ID development.

bit behind because CADD obtains documents from the Process group for drafting.

The Piping group generally does not have much involvement with P&ID development until late stages, mainly in pipe distribution drawings. The other contribution of the Piping group is placement of drain and vent valves in different locations.

The Electrical group does not have much involvement in P&ID development. They, however, need to add the required electric power of equipment on P&ID sheets.

The Civil group could end up having no involvement in the P&ID development unless the containers are of a concrete type.

Based on the previous section, it can be surmised that the workload of the Process group will decrease after the IFD version of P&IDs, while the workload of the other groups will increase.

2.4 P&ID Set Owner

In an engineering company, each document has an owner and a P&ID is no different. The owner of a document is neither the person who has the sole liability regarding the content of that document nor the sole person who uses that document. However, a person or group should be assumed as the go-to person for issues regarding that document. The ownership of a document may be changed at certain times.

A P&ID is not different than other documents; however, the owner of P&ID has a critical role due to the importance of P&ID during the design phase. It is natural tendency for companies to assume that the Process group is the owner of the P&IDs – which is not always the case. In smaller projects in which process activities are limited, the Mechanical group could be assumed to be the P&ID owner. In projects that are executed on an aggressive schedule, the P&ID owner could be the Project Engineering group. Even sometimes when the Process group is selected as the P&ID owner, the ownership of the P&ID can be transferred to the Piping group or Project Engineering group after IFD revision of P&ID. This can be done based on the logic that after IFD gate, the Process group will not be the main player of the game in P&ID development. The other reason could be, as some people say, that “if the P&ID remains in the hands of the Process group, then P&IDs will never be finished because the Process group tends to keep changing them!”

Within the Process group, the P&ID development could be handled in a nonstructured way, or it can be put on the shoulders of a P&ID administrator or a P&ID coordinator. In large projects, the responsibility of a P&ID administrator is handling the nontechnical

aspects of P&ID development and also ensuring uniformity in P&IDs.

2.5 Required Quality of the P&ID in Each Stage of Development

During the evolution of the P&ID, more details with more accuracy are shown on the P&ID. At the early stage of P&ID development, a person may not see any small-bore, 2” pipe, but in the last stage of P&ID development, IFC revision of that, all 2” pipes should be shown. In the early stage of a project, the length of a vessel could be mentioned at 5000 mm, but the same length would be seen on an IFC revision of P&ID as 5200 mm.

But how can one say an issued P&ID as IFA has been produced according to quality standards? Each company has its own standard of quality for the P&IDs in each stage of development. However, a general understanding may exist and can be used as guideline. Because we are not still familiar with *all* of the elements on a P&ID, such a guideline cannot be explained here. A guideline of acceptable quality of P&IDs is discussed in Chapter 18.

2.6 P&ID Evolution

During the development of the P&IDs, different parties cooperate to increase the quality of the P&IDs. Therefore, P&IDs evolve during the design stage of projects.

P&IDs evolve in three ways: additions, deletions, and changes. Different individuals from different groups do either or all the “P&ID development actions” on P&ID sheets in the form of “markups” every day or every few days.

From time to time, the marked-up P&ID sheets are sent for redrafting to implement all the markups and to produce a clean copy of the P&ID.

Some companies have specific rules regarding the marking up of the P&ID to make life easier for the Drafting group. A typical guideline is shown in Table 2.1.

At the early stages of P&ID development, all the additions, deletions, and changes should be done to make sure the P&ID is the best fit for the considered purposes. Later during P&ID development, only the additions, deletions, and changes that are musts, and not necessarily preferences, should be done.

2.7 Tracking Changes in P&IDs

Do we need to keep track of changes or not? How can we keep track of constant additions and changes?

Table 2.1 Change markups on P&IDs.

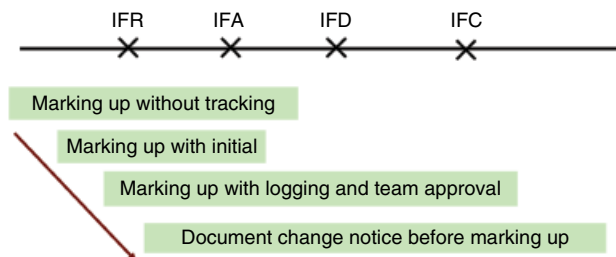
Intention	Color used
Addition	Blue
Deletion	Red
Comment	Black

At the beginning of the P&ID development, there is no need to keep track of changes as there can be huge additions or changes on P&IDs, and also those changes generally do not have any impact on other groups' schedule and budget. This period could be the pre-IFA period.

After getting close to a stage that a “skeleton” of a P&ID is developed, the program of keeping track of the markups should be started. For example, it could be instructed that every change on P&ID should be logged with initials and the date. Or the procedure could be that every single change should be marked up and initialed, but it will remain there if it is approved by the Process group or by the Process group *and* the project engineers. This type of tracking program could be implemented for the P&ID development between IFR and IFA or IFA and IFD.

After issuing an IFD begins a critical period of P&ID development. In this stage, all other groups start to do their work. Then, every change on the P&ID may impact them. Therefore, a more stringent tracking program should be implemented. In this stage, no P&ID change can be added unless it is approved and signed off by other groups including the project engineer or manager.

Therefore, for P&ID additions, deletions, or changes after IFD, a document package should be prepared and the change title logged in a log sheet by the proposer of the change for *every* change. There is usually a P&ID change meeting on a weekly or biweekly basis to discuss

**Figure 2.4** Management of change on P&IDs.

and approve or reject every proposed change. In this stage, no change on a P&ID can be marked up before approval by the group lead. In later stages of project (e.g. post-IFC stage), every single change may need to be filed in a change notice.

This concept is shown in Figure 2.4.

2.8 Required Man-Hours for the Development of P&IDs

It is difficult to introduce a methodology for predicting the required man-hours for P&ID development because each company may use its own methodologies for predicting the required man-hours. The required man-hours depends on the available tools in companies, the maturity and completeness of their P&ID development guidelines, and the skill of the designers.

The estimation of man-hours is normally the responsibility of each group. Process, Instrumentation and Control, Drafting, Piping, Mechanical, Civil, and other groups estimate their own hours for P&ID development. However, in some engineering companies, only two groups (Process and I&C) consider and assign man-hours under the category of P&ID development. The other groups generally do not assign specific separate man-hours for their P&ID involvement; it could be because their involvement is not as large as that of the two main contributors of P&ID development.

The Piping group has a specific role. On the one hand, they are mainly the “users” of P&IDs wherein one may say they do not need any assigned man-hours for P&ID development. They, however, may need to contribute to the P&ID development during detail stage of project. There are cases in which the Piping group design affects the P&ID. In such cases, the Piping group may need to incorporate changes on the P&ID.

The goal of man-hour estimation for P&ID development is coming up with required man-hours for developing each single P&ID for each incremental development or for development from the beginning to the IFC revision of P&ID. For example, a process engineering lead

may say: “for this P&ID sheet there is a need to expend 100 hours from the IFR to IFC.”

It is important to know the time required to develop a P&ID sheet comprises five time spans:

- 1) Time to develop the technical content of P&ID.
- 2) Time to get the drawing drafted.
- 3) Time required by the reviewers and approvers to check and sign the P&ID.

- 4) Time required to implement the reviewers’ comments on the P&ID.

- 5) Time that a “Document Control group” needs to officially issue the P&ID sheet.

The first item is the technical component of the man-hours, which relates to developing the technical content of P&ID.

3

Anatomy of a P&ID Sheet

Figure 3.1 shows a typical P&ID sheet composed of different blocks. Companies may decide to have different blocks on their P&IDs, but their sheets should have at least a Title block and an Ownership block. The Title block tells the reader what something is, and the Ownership block indicates who made the P&ID and for whom it was made.

3.1 Title Block

The Title block is like an ID card for a P&ID sheet. The important technical information in a Title block is the type of drawing (which in this case is a P&ID), the P&ID sheet name, number, and revision number or revision letter.

The nontechnical information in the Title block are the name of the client and the engineering company, the job or project number, and so on. The revision number is an important information. It shows the level of reliability of the P&ID; a Rev. 1 P&ID sheet is more reliable than a Rev. 0 of the same P&ID sheet. The revision number in the last row of the Revision block should match the revision number in the Title block (Figure 3.2).

It is inevitable that sometimes one or several P&ID sheets are eliminated from the set. This can be reflected on the specific, predetermined revision number of the P&ID sheet or in a large diagonally written phrase to clarify that.

A P&ID sheet could be removed because of a change in project scope or other decision that makes the content of P&ID sheet irrelevant or because the project is transferred from one company to another. A company may use the words *void*, which means its content is no longer needed in the project, and *obsolete* when a P&ID is moved to another company.

It is strongly recommended that a P&ID is never removed for minor reasons like moving the content to other P&ID sheets.

The name box of a Title block states the type of drawing, which is always a P&ID, and a specific name based on the content of a P&ID. It is a good practice to not use too specific names for P&IDs as putting more than

enough information may cause problems in the future. For example, never ever include the equipment number in the P&ID name because an equipment tag number may be changed, which then requires a change to the P&ID name, too. Some companies are very strict regarding the changing of a P&ID name: If a P&ID sheet is to be revised, the whole P&ID should be eliminated instead and a new P&ID sheet with a new number should be created. Therefore, a P&ID name like “Methanol Injection Package” is better than “Methanol Injection Package K-231.”

3.2 Ownership Block

In the Ownership block of a P&ID, the names of the owner and the designer are mentioned. If legally necessary, the credibility of the designer should be also indicated (Figure 3.3).

3.3 Reference Drawing Block

In this block all other drawing which are needed to be studied previously to have a complete understanding of the P&ID, are listed. Basically, studying the reference drawings is a prerequisite to studying the other P&ID sheets. Typical reference drawings are legend sheets and auxiliary P&IDs. Legend drawings are the most important reference drawings that should not be skipped because they introduce and define the meaning of different symbols in a P&ID set (Figure 3.4).

3.4 Revision Block

The Revision block is located at the bottom of each P&ID sheet representing the “freshness,” and hence the reliability, of the P&ID sheet. It shows revision names and their date of issue (Figure 3.5).

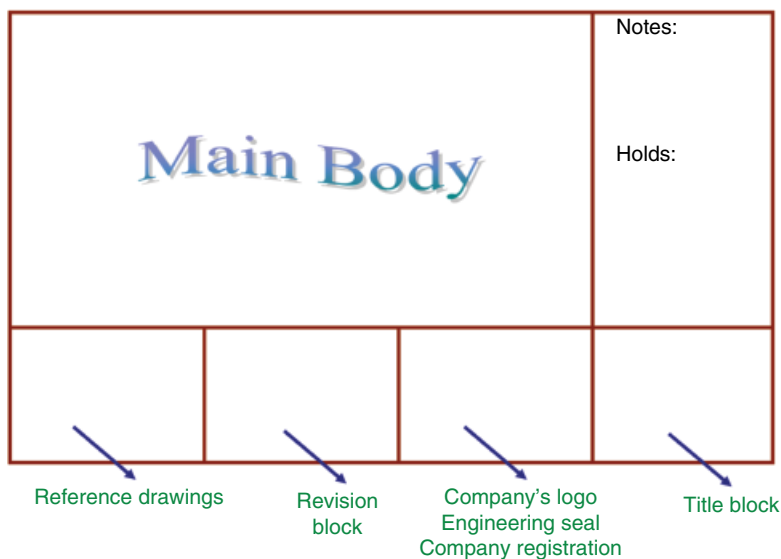


Figure 3.1 The outline of a P&ID sheet.

The Revision block basically shows the creation history of a P&ID sheet. By studying the Revision block, one can know when each revision of P&ID was issued and who are involved in the P&ID development and approval. The different revisions were discussed in Chapter 2.

Rich guys Co.		Money Maker project	
PIPING AND INSTRUMENTATION DIAGRAM FWKO Drum			
JOB NO.	1234		
DRAWING NO.			REV.
01-PID- 05-1001		0	

Figure 3.2 Typical Title block.

3.5 Comments Block

The Comments block is the area where everything that cannot be pictured in the main body of a P&ID is shown. It can have a *Holds* area and a *Notes* area (Figure 3.6).

The Holds area is not used by all companies. The general rule in P&ID development is that if an item or a portion of a P&ID does not meet the quality and completeness expected for a given revision level, it should be *clouded* and the word *hold* should be written near the border of the cloud.

Completeness in regard to the revision level of a P&ID is very important. If whatever is incomplete is supposed to be clouded, *all* the items on the first revision of a P&ID (e.g. issue for review [IFR]) should be clouded. But this is

Engineer and permit stamps
<div style="border: 1px solid black; padding: 10px; width: fit-content; margin: 10px auto;"> Permit to practice Signature:----- Date:----- Permit number:----- </div>
<p>Legal terms</p> <p>This document was prepared exclusively for XXXX by YYYY and subjects to the terms and conditions of its contract with YYYY.....</p>

Figure 3.3 Typical Ownership block.

DWG.No.	Reference drawings	REV.
52-27-002	HVAC detail DWG.	A

Figure 3.4 Typical Reference block.

not always the case. In an IFR revision of a P&ID sheet, only the items that should be clouded are those that are expected to be completed and firm in the IFR revision but are not. Therefore, to be able to decide if one item on the P&ID should be clouded, company guidelines should be followed. In the case of lack of company's guideline, Table 19.4 in Chapter 19 can be consulted.

Figure 3.5 Typical Revision block.

0	22/10/2010	Issued for approval	RP	LP	PE	CL
No.	Date	Revision	BY	ENG.	APPR.	APPR.

Notes
<ol style="list-style-type: none"> 1. Vents and drains shall be provided at the suitable locations during the construction. 2. Deleted 3. Deleted 4. PVRVs complete with bird screen.
Holds
<ol style="list-style-type: none"> 1. The sizes of all relief valves are preliminary and to be confirmed during detailed engineering stage. 2. Centrifuge tie-point for vendor confirmation.

Figure 3.6 Typical Comment block.

For companies that have a Holds area on a P&ID, the reasons for the hold can be stated here. The importance of having a Holds area is that it prevents confusion. For example, the following conversation highlights this.

Engineer 1: Oh, the pump flanges is still missing? Put the pump flanges sizes please
 Engineer 2: Two month ago I received an email from the vendor and they said the flange sizes are tentatively 6" and 4", but they never confirmed.
 Engineer 1: ok, put them here on P&ID for now.

Later during pump installation it was found that the flange size by the vendor are not correct and they wanted to firm it up, which never happened.

This problem could be prevented by putting tentative flange sizes, cloud it, and in hold area put a comment as: "To be confirmed by vendor"

The *Notes* area is where all the information that cannot be presented as symbols in the main body of a P&ID are stated. Because a P&ID is a visual document and tool, the

Notes should be eliminated or at least the number of notes should be minimized.

The Notes can be classified in different ways: main body notes and side notes; specific notes or general notes; and design notes or operation notes.

Where placing a note on a P&ID is inevitable, the best location for a note should be determined, whether in the main body or the Notes block of the P&ID. If a note is specific to an item *and* is very short, it can be placed on the main body near the item. But if a note is general or long, it is better to be placed in a Notes block.

Notes should not be more than three or four words in the main body of the P&ID. For a practical approach, main body notes are more preferable than side notes because they can be easily recognized compared with side notes that will be more likely overlooked.

Notes in on the side could be specific or general. Specific notes are those that refer to specific points of the P&ID. Such notes should have the corresponding phrase "Note X" (X represents the note number in the Notes block) in the main body of P&ID. This concept is shown in Figure 3.7.

One common problem in P&IDs are "widowed" notes, which are the notes in Notes area with no corresponding phrase in main body of the P&ID.

General notes refer to any specific area or item on the P&ID. These do not have any corresponding Note X in the main body of P&ID.

Notes can be classified as design notes or operation notes depending on their applicability. If a note is placed for the designers (piping, instrument, etc.) and for the design duration of a project, it is called a design note. If a note gives an information to operators or plant managers during the operation of the plant, it is called an operator note (Figure 3.8).

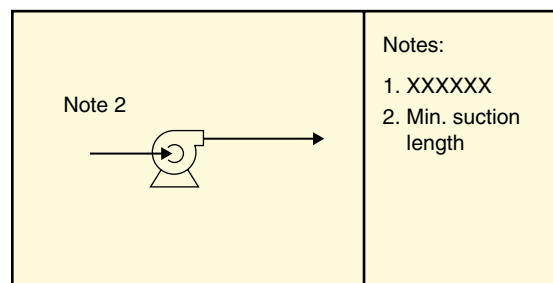


Figure 3.7 Typical Note in note side area.

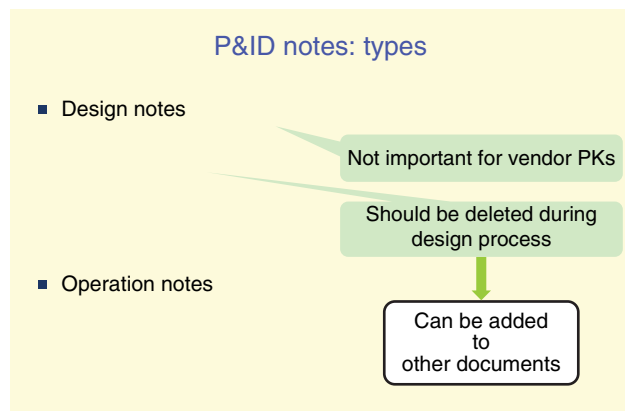


Figure 3.8 Design notes versus operation notes.

Different notes will be discussed in the next chapters relevant to each process item. Some examples of design notes and operator notes are shown in Tables 3.1 and 3.2, respectively.

Sometimes some design notes are not mentioned because other designers already know what they represent. For example, the pipe connected to the outlet flange of pressure safety valve (PSV) should be sloped against the flow (reverse slope); however, this is not always noted or mentioned on P&IDs because “pipers know this.” In

these cases, no absolute decision can be made. On one hand, it is impossible to include *all* the design notes known to every designer. On the other hand, it is best *not* to rely on other designers’ knowledge. Therefore, the design notes should be put based on the level of skillfulness of designers involved in the project.

Table 3.2 Typical operator notes.

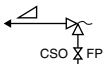
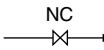
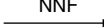
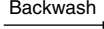


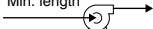

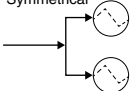
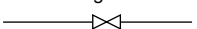

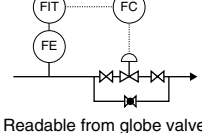

Operator note	P&ID symbol	Comment
Car seal open (CSO)		All valves in the route of pressure safety valves should be CSO to prevent mistakenly closed valve
Normally closed (NC)		During the normal operation of the plant, this valve should be closed
Normally no flow (NNF)		Intermittent stream
Backwash pipe		This stream is for backwashing
Future/spare nozzle		This nozzle is spare and for future function

Table 3.1 Typical design notes.

Design note	P&ID symbol	Comment
Do not pocket (no pocket)		Prevent creation of gas pocket in intermittent flows, easier draining when the pipe is out of operation
Min. length		The suction side of centrifugal pump should be of minimum length
Slug flow		There is a two-phase flow in the pipe and its regime is slug flow; the pipers need to design strong pipe supports to address it
Symmetrical branches		To make sure flow is divided evenly
Locate at ground level		Possibly the pipe is on a pipe rack, but it should be brought down for the valve
Vertical run		For safety reasons, the outlet of pressure safety valve; if it releases to the atmosphere, it should be vertical and upward
FIT should be in the vicinity of the globe valve (or FIT should be readable from globe valve)		For operator during the control valve maintenance: To read the flow from the flowmeter and adjust the flow by the manual globe valve accordingly
Full port (FP)		All valves in the route of pressure safety valves should be full port (or full bore) so as not to obstruct the flow

Whether to keep the design notes or not is another issue. Some companies believe that the intention of design notes is directing other designers during the design stage of a plant and that all of the design notes should be eliminated after issuing IFC version of P&IDs. However, other companies leave the design notes because they can be helpful during the operation of the plant when something needs to be changed and replaced with another design note. Also the deletion of each note – either the design note or the operator note – does not leave a clean space as the note number should be left so that the remaining note numbers still stand.

Visual notes can be used in some circumstances. They can be the detail of construction for a complicated part. They are, however, are not usually placed on P&IDs.

3.6 Main Body of a P&ID

On the main body of P&ID, the main elements that can be seen are “item identifiers,” which will be discussed in the next chapter.

There is *always* a debate about what needs to be put on the main body of the P&ID. If the P&ID is a focal document in the process plant, do *all* the information provided by all groups need to be put on the P&ID? It is NOT easy to

Lack of information Legibility Congestion




Figure 3.9 The amount of content on P&IDs.

decide what and how things should be shown on P&IDs; this topic will be expanded more in Chapter 4.

However, it should be mentioned that the amount of information shown on a P&ID can be problematic. Thus a P&ID is created “universal” for everyone involved in process plant.

If the amount of information is not enough, the P&ID becomes useless. If the amount of information shown on P&ID is too much, the P&ID also becomes useless because of its illegibility. It is important to know that a P&ID will be used in the process plant during operation, including emergency situations. An operator should be able to grasp the information needed from a P&ID as quickly as possible.

However, it is not always easy to gauge and find the sweet window for the amount of information that should be shown on a P&ID (Figure 3.9).

There was a time that there was no single “footprint” of interlock system on P&IDs. Now, a lot of P&IDs show the interlock systems, too.

4

General Rules in Drawing of P&IDs

This chapter discusses what is shown in the main body of a P&ID sheet, which is followed by a discussion of the different types and names of P&IDs based on their content.

4.1 Items on P&IDs

Anything related to the process or anything needed to present the journey of raw materials into becoming final products should be shown on a P&ID.

The above mentioned fact can answer many questions such as “Do we need to show the HVAC system of an indoor process plant?” In some cases, a portion or the whole plant could be indoors. In indoor plants, there can be an HVAC system in the building(s) to create a more suitable atmosphere for operators and equipment. As a general rule, very few details of an HVAC system are shown in such plants. However, in HVAC industries, the P&IDs can be drawn with their main purpose, that is, adjusting the air parameters.

There are basically four different items that can be shown on P&IDs:

- 1) Pipes and other flow conductors.
- 2) Equipment.
- 3) Instruments.
- 4) (Instrument and control) signals.

4.1.1 Pipes or Other Flow Conductors

Pipes and other flow conductors such as pipes, trenches, channels, and so on direct and transfer fluid from one equipment to another. The general rule is that the flow conductors of the main process fluids should be shown in the P&IDs along with the pipes. In a water treatment plant, the water flows in channels, so the channels are shown, too.

One important exception are tubes, which are not generally shown in P&IDs. However, there can be some

“footprints” of tubes that can be seen on P&IDs. This will be discussed in chapters 13 and 18.

The items for transferring bulk materials are generally categorized as “equipment” rather than “flow conductors.”

When it comes to showing pipe fittings, there is one rule: No pipe fittings are shown *except* tees, reducers, process flanges, and cap, plug, and blind flanges (Figure 4.1).

A straight piece of pipe on a P&ID could be a pipe circuit in-field with a bunch of elbows. A straight piece of pipe in-field can be represented as a line with several directional changes on a P&ID.

4.1.2 Equipment

The main players in processes are the equipment such as pumps, compressors, heat exchangers, and reactors. Containers can arguably be classified as equipment, too. Tanks and vessels are for process and/or storage purposes. All equipment should be shown on the P&IDs. If the equipment, however, are purely associated with mechanical details that are not related to the process, they may not be shown on the P&IDs. Examples are a gear box associated with a mixer, small built-in lubrication systems, and power hydraulic systems. In large compressors, the lubrication system can be large and a separate system. In such cases, the lubrication systems are shown, too.

Equipment can be metallic, fiberglass, concrete, and so on, and in all cases, these should be presented.

4.1.3 Instruments

To implement every process, two requirements should be met: the process element (i.e. equipment) should be designed and tailored for a certain process and the control system should be formed to ensure implementation. If one of these is not followed, it is most likely that the process goal will be only on paper.

Instruments are the hardware that implement the control strategies in the plant. Industry practices with

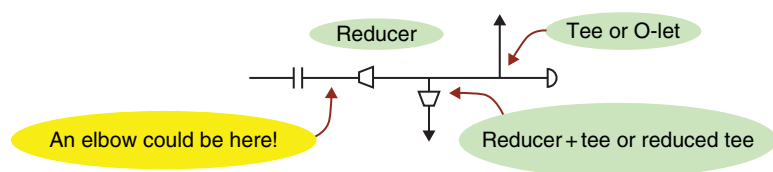


Figure 4.1 P&ID representation of pipe circuits.

Table 4.1 Items by different groups on P&ID.

Group	Shown	Not Shown
Process	Few operational details	Process discipline by itself does not have any element to show.
Mechanical	All the equipment in process or “close” to process	Mechanical-related and not Process-related items
Instrumentation and Control	Sensors, transmitters, controllers, Automatic valves	Signal processing elements, untouchable elements
Piping	All pipes, fittings, valves	Tubes
Civil	Building border, ground border, secondary containment area, sea-level concrete equipment	The rest
Electrical	Electric motors, electrical heat tracers as symbol	The rest

regard to instruments are not simple and will be discussed in Chapter 13.

4.1.4 Signals

Signals provide the means for instruments to communicate with each other. After deciding whether an instrument should be shown, the rule for showing signals would be easy: if there are two separate instruments and if there is a signal between them, this signal should be depicted.

Table 4.1 outlines some guidelines on the items generally shown on the P&IDs for each group.

4.2 How to Show Them: Visual Rules

The first thing that should be answered in regard to the P&ID is what goes where and how it is displayed.

The first thing that a reader of a P&ID expects is legibility. A P&ID set is developed by the engineers during the design step of the project but will later be used in a process plant during operations. It is used by individuals with different levels of knowledge on and familiarity with a P&ID. A P&ID may need to be read by engineers, managers with management degrees, trade practitioners, and many others. It also should give enough information about the project that readers from different backgrounds will be able to understand it. A P&ID can be used during normal operations or an emergency event. Therefore, a P&ID should be

as legible as possible. Below are some rules of thumb regarding the visual aspects of P&IDs:

- 1) The P&ID sheet is almost always in landscape orientation (Figure 4.2).
- 2) Limit the number of main equipment shown on each sheet. A crowded P&ID should be avoided. Some companies try to limit the number of items on each P&ID to four or five, but other companies allow as many as 10.
- 3) A P&ID is a pictorial document, so minimize notes on the P&IDs (no “note-stuffed” drawings).
- 4) Draw P&ID symbols as similar as possible to what an item looks like in reality and approximate to relative size, but remember, a P&ID is *not* drawn to scale (Figure 4.3).
- 5) Do *not* represent the real length of pipes on P&IDs. P&ID is a “Not To Scale” (NTS) drawing. Therefore, a short line on a P&ID could represent a few hundred meters of pipe.
- 6) Do *not* “pack” symbols on one side of the P&ID sheet. The symbols should be fairly spread out.
- 7) Generally, an equipment is arranged horizontally on one level. On an ideal P&ID sheet, any imaginary vertical line should, at maximum, cross *one* equipment symbol. The P&ID will be too difficult to understand when equipment symbols are stacked on each other and the connection between the symbols and the equipment callouts are difficult to be identified. (Figure 4.4).
- 8) All the different elements on a P&ID sheet should be connected to each other. If a group of elements have

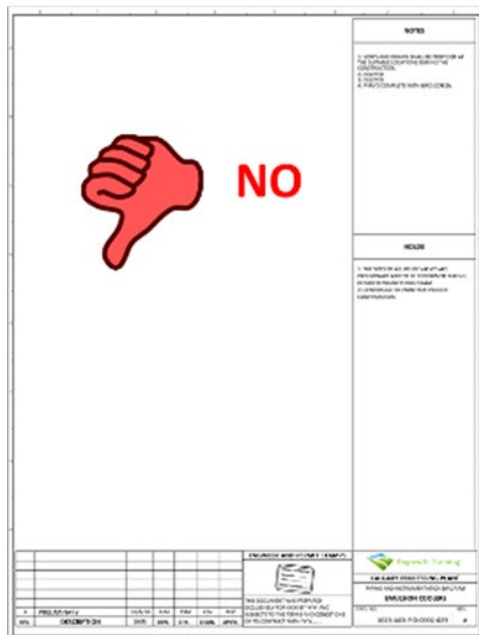


Figure 4.2 P&ID orientation.

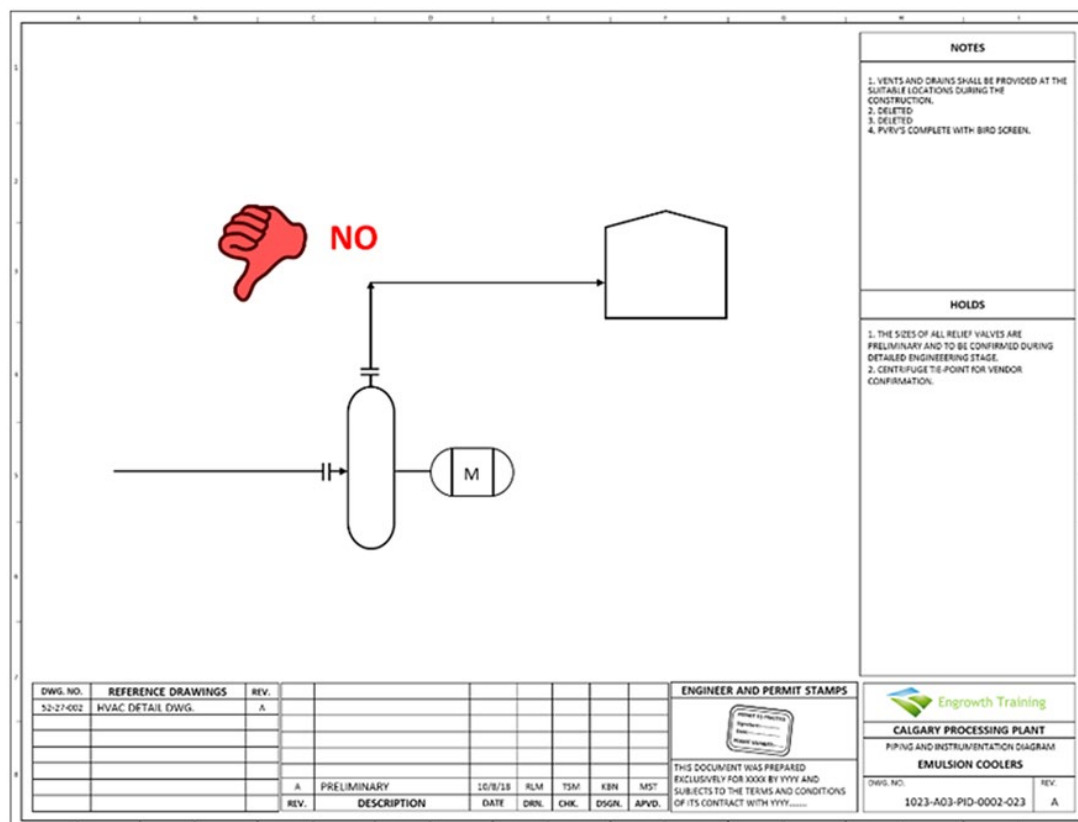
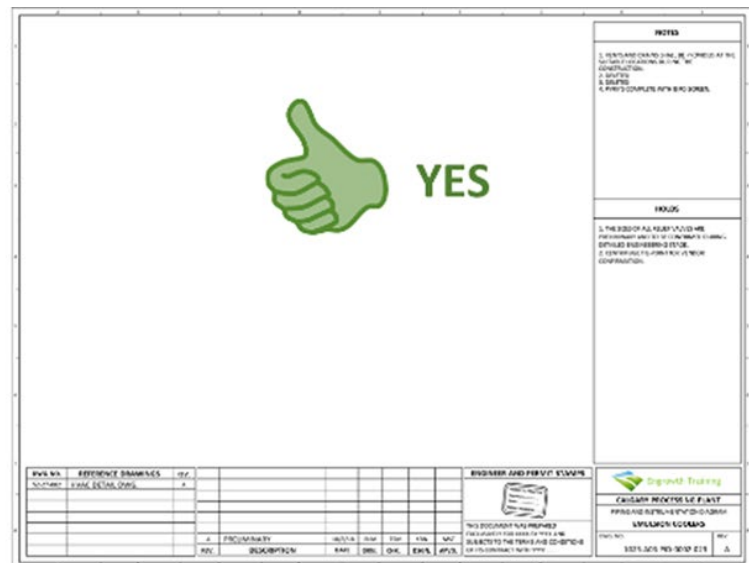


Figure 4.3 Keeping the relative size of equipment on the P&IDs.



Figure 4.4 Equipment should be fairly distributed horizontally on the P&ID.

no connection to the rest of elements, they can be drawn on another P&ID sheet.

- 9) No visiting stream is allowed. A visiting stream is a stream that has no connection with other items on the P&ID. See Figure 4.5.
- 10) Do *not* try to present a P&ID in a way that follows geographical directions (e.g. north). P&IDs are drawings independent of these.

The rest of this section outlines the other visual issues in drawing P&IDs and the general practices in dealing with these issues.

4.2.1 Line Crossing Over

Lines (in different forms) are symbols for pipes or signals. Crossing lines should be avoided or kept to a minimum, as well as changing the direction of lines. However, sometimes both are inevitable, and it is common to see lines crossing each other in P&IDs.

It is important to know that crossing lines in a P&ID does *not* reflect the reality of pipe routes in field. Crossing lines in the P&IDs is not acceptable aesthetically, but it is



Figure 4.5 Showing visiting streams is not a good practice.

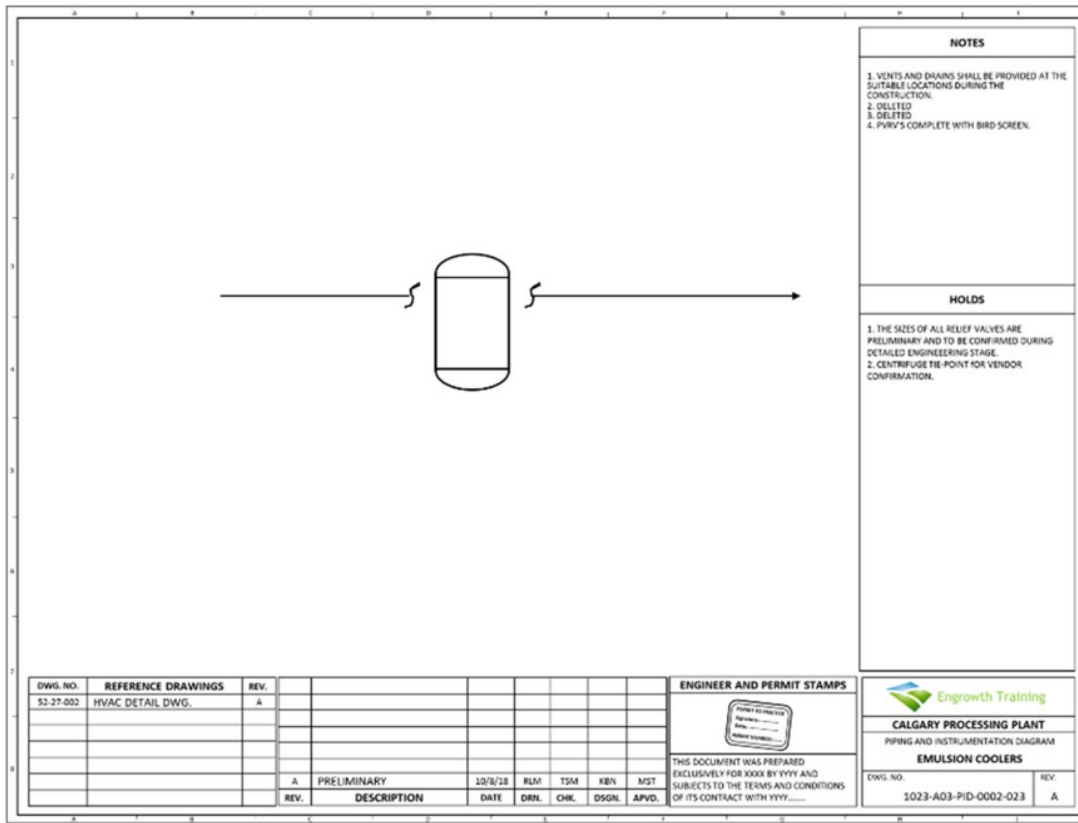


Figure 4.10 Equipment–line crossing is allowed.

Meanwhile, equipment–line crossing can be shown, but it is not a good practice. This can be managed by using breaks on the line (Figure 4.10).

4.2.3 Off-Page Connector

Off-page connectors are continuation indicators for the lines, which are pipe and signal symbols. The preference is to show off-page connectors horizontally and at the edge of a P&ID sheet. The incoming off-page connectors are preferably located at the left edge of the P&ID sheet, and the outgoing off-page connectors at the right edge of the P&ID sheet.

There are, however, exceptions.

The off-page connectors in network P&IDs (this will be discussed later in this chapter) can be drawn in vertical positions, too. This is because network P&IDs should generally follow the plot plans, and sometimes using vertical off-page connectors prevent crowding on the P&ID.

In some system P&IDs (discussed in section 4.4.2), the off-page connectors of utility pipes are shown near the utility user and not at the edge of the P&ID sheet. Some companies adopted this practice to decrease crowding on the P&ID by avoiding drawing of long utility lines, which are not as important as process lines.

The incoming off-page connectors can be on the right edge of the P&ID sheet if they are coming from a downstream of the process. The outgoing off-page connectors could be on the left edge of the P&ID sheet if they are going somewhere on the upstream of the process. The different appearances of off-page connectors are shown in Figure 4.11.

4.2.4 Color in P&IDs

Color in P&IDs is a recent progress. After the color printers gained popularity, P&ID developers decided to use color P&IDs to make them easier and quick to understand. Different colors are mainly used for pipes and signals because they primarily fill the P&ID sheets.

4.3 Item Identifiers in P&IDs

For each item to be shown on a P&ID, there can be a maximum of four different identifiers. This is like individuals in a society; each individual can be identified by different identifiers including name, ID number, picture or symbol, and brief information like eye color, hair color, height, and so on. Table 4.2 shows this analogy.

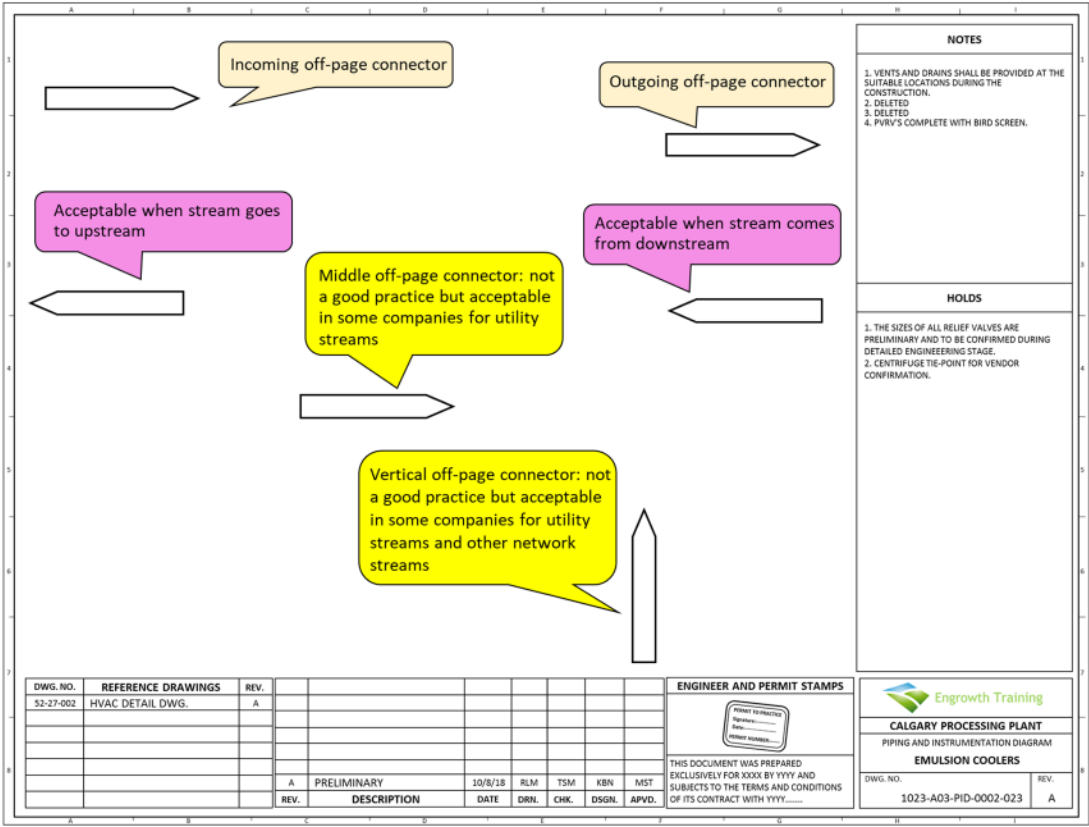


Figure 4.11 Off-page connector appearance.

Table 4.2 Analogy between a man identifier and a P&ID components identifiers.

Analogy: Individual in a society	Process element on a P&ID sheet

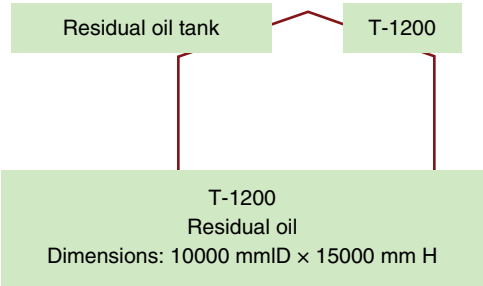


Figure 4.12 A single element on the P&ID.

Table 4.4 outlines the identifiers' location on P&ID main body.

Figure 4.12 shows a tank with all four identifiers. However, neither all items on a P&ID have all identifiers nor their identifiers are separated from them. For example, a manual valve may only have symbol and no other identifier. In the pipe tag, some technical information can be seen, too.

To be able to draw a P&ID, we need to know which identifiers are used for each item and where they can be found. Table 4.3 shows which identifiers are used by telling which identifier is provided by what group for the P&ID.

4.3.1 Symbols

Symbols are always based on the P&ID legend sheet, but the legend sheet is not always the first P&ID sheet developed, so it may be difficult for designers to pick a symbol when there are no legends to refer to. This issue, however, is not always detrimental. Process engineers generally have some previous experience and also use symbols that can be understood using common sense. However, there can be an issue, especially if

Table 4.3 Source of information for elements' identifiers.

	Equipment	I&C Items	Piping	Signals
Symbol	Per P&ID legend	"Balloons" per P&ID legend sheet (possibly based on ISA legend)	Per P&ID legend sheet	Per P&ID legend sheet (possibly based on ISA legend) ^a
Tag	From equipment list By Mechanical Engineering group	From instrument list (automatic valve list) By I&C group	From the Line List (or LDT) By Piping group	N/A
Name	Process group decision	N/A	N/A (unless on off-page connectors)	N/A (unless on off-page connectors)
Technical information	In the form of callout Decision on components per project decision, data is provided by different groups	Per I&C group's decision (Process group may put preliminary sizes for control valves and safety valves)	For pipes, part of their tag, can be for fittings and their sizes by Process group	N/A

^aI&C, Instrumentation and Control; ISA, International Society of Automation; LDT, Line Designation Table.

Table 4.4 Location of different identifiers for different P&ID items.

	Equipment	I&C Items	Piping	Signals
Symbol	In the main body of P&ID			
Tag	Close to the equipment symbol, but if there is enough room, inside of the symbol	Inside of symbol	Close to the symbol	N/A
Name	Below the tag	N/A	N/A	N/A
Technical information	Callout: On the top on bottom of main body	Close to the symbol	N/A for pipes, For appurtenances: close to the symbol	N/A

I&C, instrumentation and control.

there are different groups working on different areas of a project. One group may choose different symbols for a single element. Therefore, communication plays a vital role here.

Equipment symbols may have internals that may or may not be shown. If shown, they should be a dashed line.

4.3.2 Tags

Tags are issued by the group responsible for buying a given item. Whoever tags an item needs to buy the item. For example, an equipment tag is issued by the Mechanical group because they are designated to buy the equipment for the project. This is similar with piping tags that are issued by the Piping group.

Tags generally consist of letters and numbers in a specific arrangement (anatomy). The anatomy of instrumentation and control (I&C) items are mostly universal and based on International Society of Automation (ISA) standards. But the tag anatomy for equipment, pipe, and pipe appurtenances are defined by the client or project.

For example, in a company, the tag anatomy for equipment could be a representative letter (or block of letters) and a number, which is a sequence number taken from a table. When a mechanical engineer wants to tag a tank, he/she consults with the mechanical procedure document to see what the acronym is for a tank. It could be T or TK. The tank tag will be T-xx or TK-xx. The xx here is a number and the engineer will pick up the number from their equipment list. Therefore, the tank tag could be something like T-231.

The tagging of parallel similar equipment is an interesting issue. One approach is tagging them in a way that they have the same tag number; the only difference is a letter at the end of tag number. In this approach, two parallel similar pumps can be tagged as P-125A and P-125B. Another approach is tagging them like two unrelated pumps, such as P-125 and P-126. The current trend is the second approach as the first approach can be confusing during an emergency and when complete and clear instructions cannot be heard by the operators. For example, the instruction "Shut down P-125A" may be heard as "Shut down P-125B" in a noisy environment.

4.3.3 Name

Name is something used only for equipment, and generally the process engineer names the equipment. Some equipment names are assigned in the early stage of a project when the process flow diagram (PFD) is developed, but some of the names should be assigned during the P&ID development stage.

There is no specific rule for the naming of equipment, unless it should be unique for each equipment within each plant to prevent confusion.

The following guidelines are helpful when deciding about the name for an equipment:

- The type of the equipment is part of the equipment and generally the last word, for example, the word *pump* in a boiler feed water pump or the word *tank* in sales oil tank.
- The equipment names are rarely longer than five words. An acronym can be used if necessary, for example, BFW pump rather than boiler feed water pump.
- The equipment names are chosen generally based on their service (e.g. sales oil tank) or their position (e.g. flotation vessel discharge pump) or both (e.g. disposal water injection pump).
- Where there are several heat exchangers in a series, the last one (which is likely a utility heat exchanger) can be named with *trim* at the beginning.
- Where there are two pumps in a series to inject a certain amount of liquid to the process, the word *rough* for upstream pump and *trim* for the downstream can be used at the beginning of their names.
- Where there are two pumps in a series and the upstream pump is used to provide enough net positive suction head (NPSH) for the downstream pump, the word *booster* can be used at the beginning of its name for an upstream pump.
- Where a fluid mover takes suction from an important equipment (like a unit operation or process unit) and is on the downstream of the equipment, it can be named as an equipment suction fluid mover like “reactor discharge pump.”
- Where a fluid mover feeds an important equipment (like unit operation or process unit) and is on the upstream of the equipment, it can be named as an equipment feed fluid mover like “reactor feed pump.”

4.3.4 Technical Information

Technical information is information about an item on the P&ID. The number and nature of information required to be placed for each item can be decided by the client. However Table 4.5 can be used in lack of better information. It is important to consider that the amount of technical information should be kept at a minimum on the P&IDs. The P&ID is a pictorial document, and

Table 4.5 Type of technical information for different items.

	Technical information
Equipment	In the form of callout; refer to text for explanation
I&C items	Sensors: range Automatic valves: size (and pressure rating) Pressure safety valve: set point, size (including inlet and outlet size and orifice designation letter)
Piping	Pipes: as part of their tag, pipe size and material code Manual valves: size (and pressure rating) Nozzles: size (and pressure rating) Fittings: their main sizes
Signals	N/A

I&C, instrumentation and control.

technical data should only be there to aid in the understanding of the symbols. If more numerical information about an item on a P&ID is needed, data sheets and other respective documents need to be consulted.

One general rule that applies to putting the sizes for pipes, fittings, and valves is that if their sizes can be recognized from the adjacent items on a P&ID, there is no need to show their size. For example, some companies instruct their P&ID designers to not put the size of reducers because it can easily be figured out from the attached pipes and their tags. This is shown in Figure 4.13.

Here, we focus only on equipment callouts as they need more explanations. Equipment callouts detail information about a piece of equipment up to roughly less than 10 short lines. Equipment callouts can be introduced in the general form of a Tag + Name + Additional Data. The information contained in equipment callouts is collected from different groups. However, the type of information required to be placed in callouts can be decided by the client. Equipment callouts are shown above each piece of equipment and on the top edge of the P&ID sheet.

Some companies use another approach for positioning the equipment callout on a P&ID. In this method, the callouts of static equipment should be placed on the top of a P&ID sheet, whereas the callouts of rotary equipment should be placed on the bottom of a P&ID sheet (Figure 4.14).

The anatomy of a callout is not arbitrary and is specified in guidelines by the client or engineering company. A typical equipment callout may have the structure shown in Figure 4.15.

Generally speaking, all equipment callout starts with the equipment tag and equipment name in one line. The next line of equipment callouts enumerates one to three mechanical characteristics of the equipment. These mechanical characteristics should be measurable parameters such as the diameter and height for tanks, flow rate

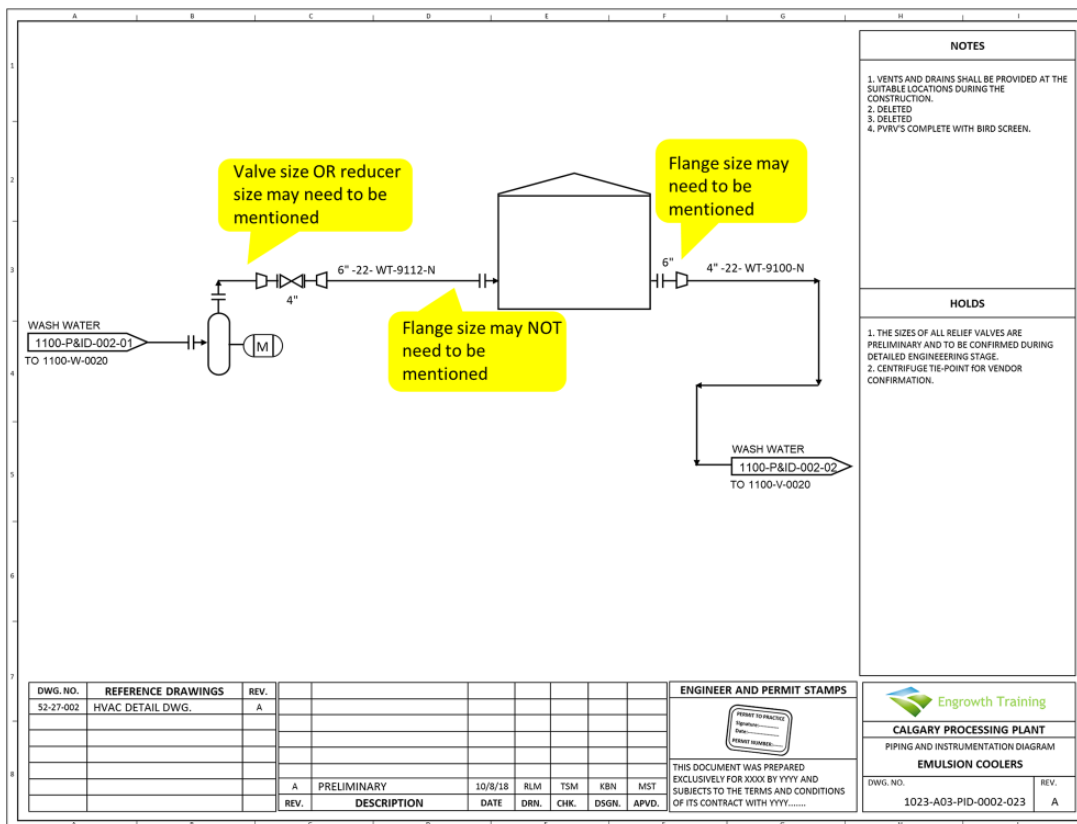


Figure 4.13 The concept of “no repetition” on a sheet of a P&ID.

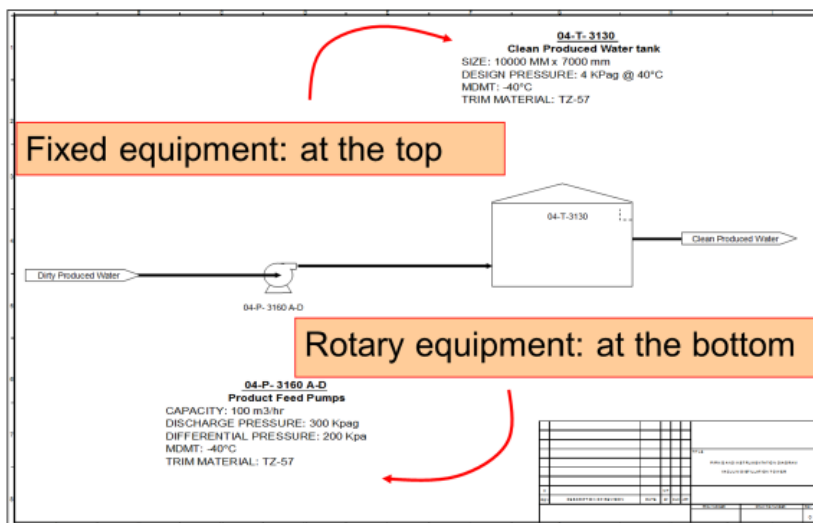


Figure 4.14 Positioning of callout for equipment.

and head for pumps, or heat transfer area for heat exchangers. A person with minimum or no knowledge about the process of the equipment should be able to measure the parameter(s).

For example, the parameters on a P&ID for a tank may be diameter and height or maximum volume (capacity), but not the normal capacity. For a heat exchanger, the parameter listed may be the heat transfer area (which

can be measured by estimating all tubes' peripheral areas in a shell and tube heat exchanger), but not heat duty, which is the heat transfer value during the operation.

The next line in the callout box is the sparing policy or philosophy, which can be explained using the concept of spare tires for cars. For example, you may have a total of five tires for a car; four of them are on the car, and one is a spare in the trunk.

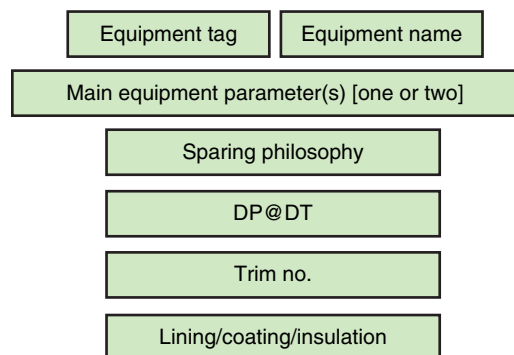


Figure 4.15 The anatomy of a callout.

So, the sparing philosophy for tires for the car in this example is $5 \times 25\%$; this means four of them are operating tires and one of them is a spare.

How is $5 \times 25\%$ interpreted as four operating ties and one spare tire?

$$\frac{100}{25} = 4 \quad 4 \text{ operating tires}$$

$$5 - 4 = 1 \quad 1 \text{ spare tire}$$

The 100 in the calculation is a fixed number.

In process plants, the sparing philosophy is a parameter that is mainly defined as a need for parallel units. For single equipment, the sparing philosophy is simply $1 \times 100\%$ and possibly does not need to be mentioned on the P&IDs. The other examples of sparing philosophy is shown in Table 4.6.

There are some variations of expressing a sparing philosophy. For example, there can be three sand filters in parallel that all function during normal operation, but when one of them is out of service and backwashing is

Table 4.6 Examples of sparing philosophies.

Sparing philosophy notation	Sparing philosophy meaning	
$1 \times 100\%$	$\frac{100}{100} = 1$	1 Operating
	$1 - 1 = 0$	No Spare
$2 \times 100\%$	$\frac{100}{100} = 1$	1 Operating
	$2 - 1 = 1$	1 Spare
$3 \times 50\%$	$\frac{100}{50} = 2$	2 Operating
	$3 - 2 = 1$	1 Spare
$4 \times 33\%$	$\frac{100}{33} = 3$	4 Operating
	$4 - 3 = 1$	1 Spare
$5 \times 25\%$	$\frac{100}{25} = 4$	4 Operating
	$5 - 4 = 1$	1 Spare

applied, just two of them are doing the filtration. This sparing philosophy can be shown in Figure 4.16.

However, when the sparing approach is more complicated, it could be more difficult to show it as a simple block of a number of parallel equipment multiplied into a percentage value. More complicated type of sparing schemes will be discussed in Chapter 5.

Design pressure and design temperature are the next items in an equipment callout. The design pressure is the pressure with which the container or a piece of equipment can operate continuously with no safety hazard.

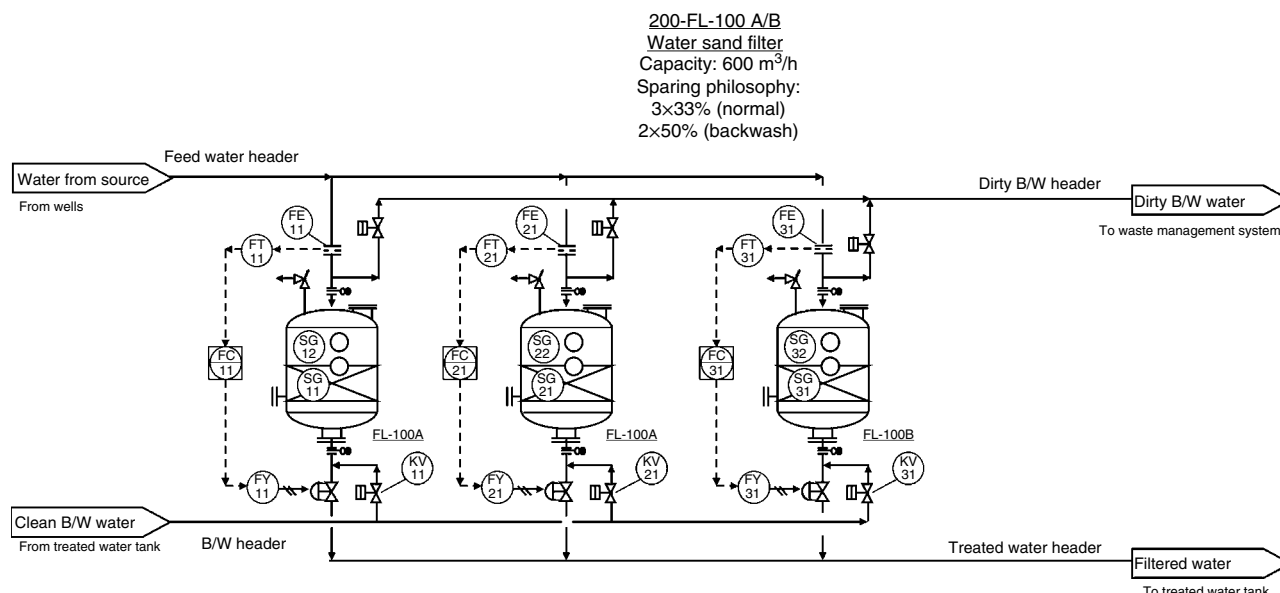


Figure 4.16 Sparing philosophy of three filters during normal operation and only two filtering during backwash.

Stating the design pressure without talking about the design temperature is meaningless. Each design pressure should be stated at a specific design temperature because generally at higher temperatures, construction materials lose their strength.

The pair of design pressure and design temperature is generally defined for containers and the casing of equipment.

It is important to know that a piece of equipment keeps its structural integrity up to its design pressure and design temperature, but it does not necessarily function properly at the pair of pressure and temperature.

The trim tag can be a confusing term in callouts. “Trim” in this context means all short pipes are connected to a piece of equipment. For example, all container nozzles have a short portion of pipe. A pump has a drain valve and a vent valve connected to the pump casing through short pieces of piping. These pipes need tags, and their tags are listed in the equipment callout. The structure of a trim tag is very similar to a piping tag. The trim number basically specifies the construction material for these little pipes connected to a piece of equipment. It is important to know that the construction material for these short pipes is not necessarily the same with that of the equipment. We may have more than one trim number in an equipment callout. These short pieces of pipes can be inlet or outlet nozzles in containers, drainage vents in pumps, and other equipment.

It is good to note that equipment and the connecting pipes were already tagged; the only “missing” tagging is for the short pipes connected to the equipment which now is covered by the concept of “trim”.

The last line of a callout can be a brief introduction about layers around the equipment. The layer can be lining, coating, or insulation for the equipment. Linings are a protective layer on the internal surface of a piece of equipment for protection against the impact of the fluid. Coating is a protective layer on the external piece of equipment for protection against the impact of weathering on the external side of the equipment walls. Insulation is applied for different reasons including heat loss abatement or sound protection.

For example, Figure 4.17 is a callout of a vessel that contains:

- Vessel tag provided by the Mechanical group.
- Vessel name provided by the Process group.
- Vessel size provided by the Process group and confirmed by the Mechanical group.
- Design temperature and pressure of the vessel provided by the Process group and confirmed by the Mechanical group.
- Minimum design metal temperature (MDMT) is requested by Process group and will be provided by the Mechanical group. (The MDMT will be discussed later.)

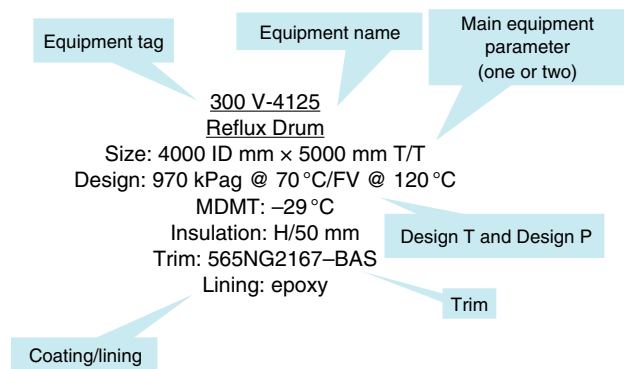


Figure 4.17 Example of a callout for a vessel.

- Features of the insulation wrapped around the vessel provided by the Process group (the insulation feature will be discussed later).
- The trim tag provided by the Process group based on the piping specifications document prepared by the Piping group. (The trim tag will be discussed later.)

4.4 Different Types of P&IDs

There are primarily five different types of P&IDs: legend P&IDs, system P&IDs, network P&IDs, interarea P&IDs, and detail P&IDs. Although not all companies recognize the different types of P&IDs, the concept exists and needs to be understood.

The legend P&IDs are a few P&ID sheets that introduce all the symbols and acronyms and their meaning. To be able to read and understand the P&IDs, there needs to be one or more sheets showing the meaning of symbols and acronyms on the P&IDs.

The core of P&IDs, system P&IDs are the P&IDs that show the material changes.

The network P&IDs are the P&IDs that show the distribution or collection of material within the plant units, for example, utility distribution (and collection) P&IDs, safety release gathering network P&IDs, fire water P&IDs, sewer collection system P&IDs, and blanket gas and vapor recovery unit (VRU) P&IDs. Some companies use other names for utility P&IDs such as utility distribution diagrams (UDD) or U&IDs. It is easy to determine if a P&ID is a network P&IDs – these are mainly pipes and does *not* have equipment.

Interarea P&IDs show where pipes or signals pass from one area to another area. There are two types of interarea P&IDs: interconnecting P&ID and battery limit (B/L) P&ID. The former shows the passing internal borders of a plant through different units of the plant, and the latter shows the passing of the plant border (i.e. plant B/L). The other name of a B/L P&ID is “tie-point P&ID.”

Interarea P&IDs may have some features of network P&IDs, but the pipes shown on them are straight and do not have a spider web appearance.

The last type of P&IDs is detail P&IDs. These P&IDs show details of some systems that are not related to the main process and do not need to be shown on process P&IDs.

The term *auxiliary P&IDs* can be used differently in each company. Some use it to refer to a detail drawing, while others use it when referring to the detail P&IDs and legend sheets or even vendor or licensors drawings. Figure 4.18 summarizes the classifications of P&IDs.

4.4.1 Legend P&IDs

Generally, the first few sheets of a set of P&IDs contain the list of all P&IDs and a legend sheet. Similar to a table of contents of a book, this list enumerates all the P&IDs in the set with their tags and their descriptions. The other few sheets are legend sheets where all the symbols and acronyms used in a set of P&IDs are explained. A sample P&ID legend sheet is shown in Figure 4.19.

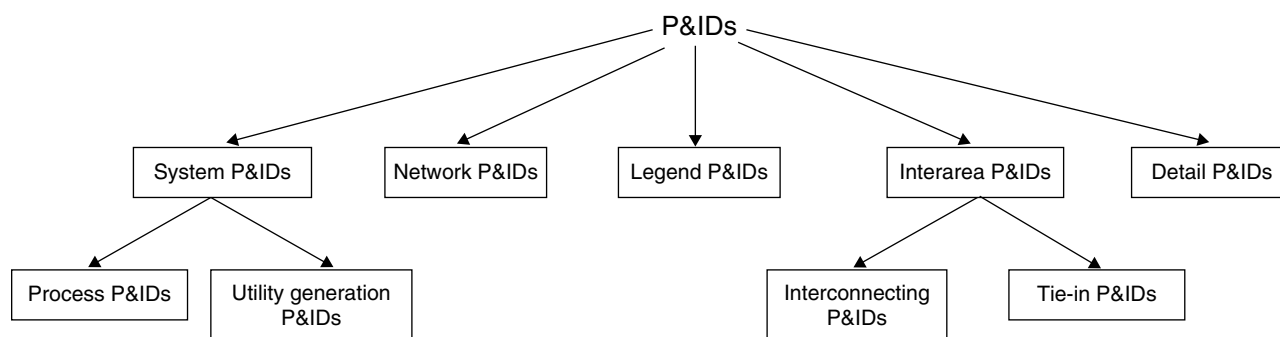


Figure 4.18 P&ID type classification.

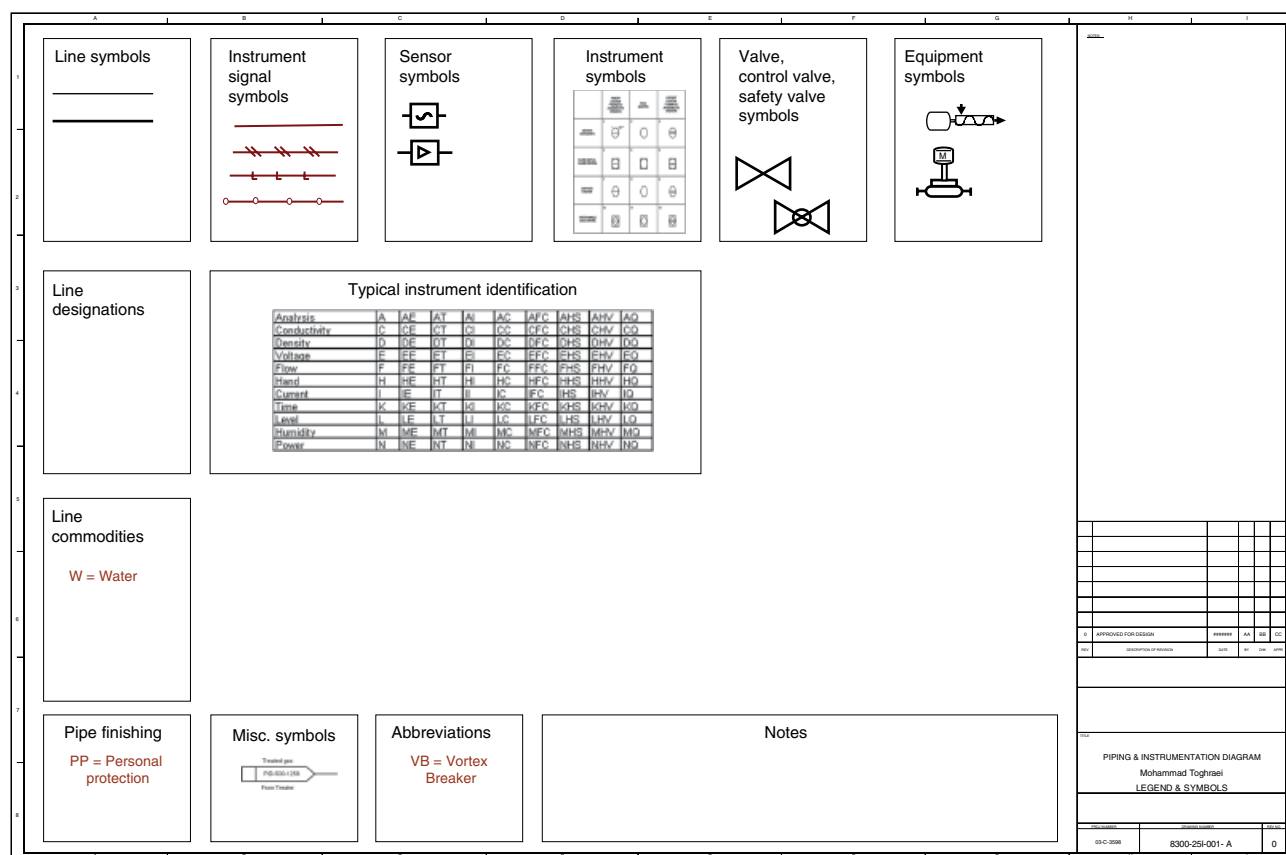


Figure 4.19 A P&ID legend sheet.

4.4.2 System P&IDs

The system P&IDs consist of two sets of P&IDs: process and utility generation or utility processing. Process P&IDs show the main process that the plant is built (or will be built) for. They are, in fact, the P&IDs of the main process plant. Process P&IDs are the most unique type of P&ID in a set of P&IDs. For example, network P&IDs or interarea P&IDs look fairly similar in a refinery P&ID set and wastewater treatment plant set, but their process P&IDs are totally different.

Utility generation P&IDs show the generation of different utilities, like steam and instrument air, in the plant. They are sort of process P&IDs, but the products are not the final product of the plant. In other words, utility generation P&IDs are essentially the same with process P&IDs but they deal with utilities and not process products. P&IDs that show a wastewater treatment plant can be classified as utility-processing P&IDs.

A sample process and a utility generation P&ID are shown in Figures 4.20 and 4.21, respectively.

4.4.3 Network P&IDs

Network P&IDs are classified based on different criteria. Network P&ID's could be classified as distribution networks or collection network. They can also be categorized as utility and nonutility networks or as continuous flow networks or intermittent flow networks. These classifications are shown in Table 4.7.

The network pipes can be designed and developed as under- or aboveground pipe networks. The service that a network functions affect its design and its P&ID development. This will be discussed more in Chapter 16.

The network P&IDs are generally drawn to follow the plot plan. This means that they can be roughly superimposed on plot plans. If there is a need to show any equipment symbol on the network P&ID, it can be shown in plan view, rather than the more common side view.

The Piping group is highly involved in developing network P&IDs. During the plot plan design, different pipe routes are implemented in the plant that allow pipes to go through them rather than at random routes for each single pipe. It is like when you want to go from your house to your workplace, there is a specific route you need to take, but you cannot make a diagonal path on the streets and cannot go over buildings. Similar to cities, there are some dedicated routes for pipes in each plant. These routes can be underground routes, pipe galleries or trenches, aboveground (like pipes on "sleepers"), or elevated pipe routes (pipe racks).

A large portion of network P&IDs are utility P&IDs. The utility distribution or collection P&IDs can be shown as a single-commodity P&ID or as a multiple-commodity

P&ID. In a single-commodity utility P&ID, on each P&ID sheet, only one utility (e.g. steam) is seen, whereas in a multiple-commodity utility P&ID, more than one utilities (e.g. steam and plant air). The tendency is toward developing single-commodity P&IDs when the plant is huge and the utility users are large. A sample utility distribution P&ID, single-commodity type, is shown in Figure 4.22.

4.4.4 Interarea P&IDs

As was mentioned, interarea P&IDs are the B/L P&ID and interconnecting P&ID. A B/L P&ID is a snapshot of different sections of the plant's borders. It shows the short portions of the pipes that cross the borders and that are seen on both sides of the border.

A B/L P&ID shows all the pipes coming from outside of the B/L (outside the plant's property) that are connected to the pipes inside the B/L. These pipes generally transfer raw material from a raw material producer or from natural resources (like oil reservoirs) to a plant, or they transfer the products from the plant to the product handling company or end user. Therefore, B/L P&IDs exist for at least two locations: at the beginning of the plant and at the end of a plant.

Generally, there is no equipment on B/L P&IDs. It is common to see the borderline and the isolation arrangement wherever the pipes cross the borderline on these P&IDs. The isolation arrangement will be discussed later; but briefly, it consists of one of the following arrangements: blind, isolation valve and blind; isolation valve, blind, and check valve.

On B/L P&IDs, tie points may be shown, too, and are therefore, known as tie point P&IDs. Tie points refer to pipes that should be connected to each other at the B/L border through mating flanges. To prevent confusion during construction, the tie points can be tagged.

In some B/L P&IDs, the ownership of materials changes. For example, a flow of raw material is transferred from company A to company B under some financial arrangement. In such cases, the flow rate or the quality of the stream needs to be measured precisely. A process package may be installed to achieve this goal; which could be known as a lease automatic custody transfer (LACT) unit. In nontechnical language, a LACT works similar to a counter at retail stores, where the quality of a product is checked by the buyer before the purchase is completed. A sample B/L P&ID is shown in Figure 4.23.

The interconnecting P&IDs show snapshots of the pipes after the B/L P&IDs. They show the distribution or collection of process streams (and not utility streams) to different areas of a plant. Interconnecting P&IDs show the pipes' routes from the B/L to their different destinations; they are



Figure 4.20 A Process P&ID.

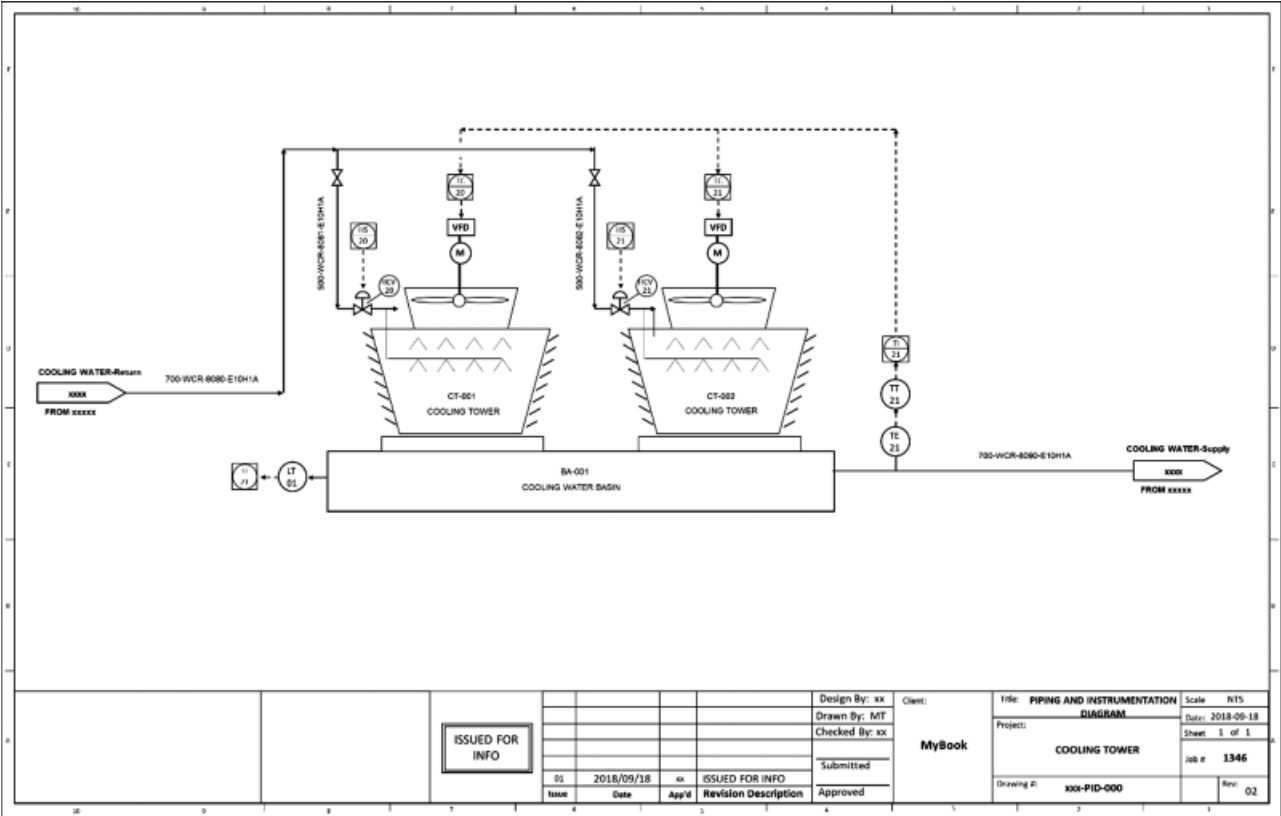


Figure 4.21 A Utility Generation P&ID.

Table 4.7 Different services of network P&IDs.

	Distribution or collection network	Utility or nonutility network	Continuous flow or intermittent flow
Utility distribution P&IDs	Distribution	Utility	Continuous flow or intermittent flow
Utility collection P&IDs	Collection	Utility	Continuous flow or intermittent flow
Relief and blowdown P&IDs	Collection	Nonutility	Intermittent flow
Fire water distribution P&IDs	Distribution	Can be considered as utility or nonutility	Intermittent flow
Sewer collection system P&IDs	Collection	Utility	Intermittent flow
Blanket gas distribution P&IDs	Distribution	Utility	Intermittent flow but very frequent
Vapor recovery unit P&IDs	Collection	Utility	Intermittent flow but very frequent

also based on the plot plan of the plant. A typical interconnecting P&ID sheet is shown in Figure 4.24.

4.4.5 Detail P&IDs

The last part of a set of P&IDs is the detail P&IDs. When some items are removed from the main P&IDs

for the sake of legibility and put on other P&IDs, those are the detail P&IDs. Therefore, each sheet of the detail P&ID should be referred from one or more main P&IDs.

The systems shown on detail P&IDs comprise systems that have any of the following features or a combination:

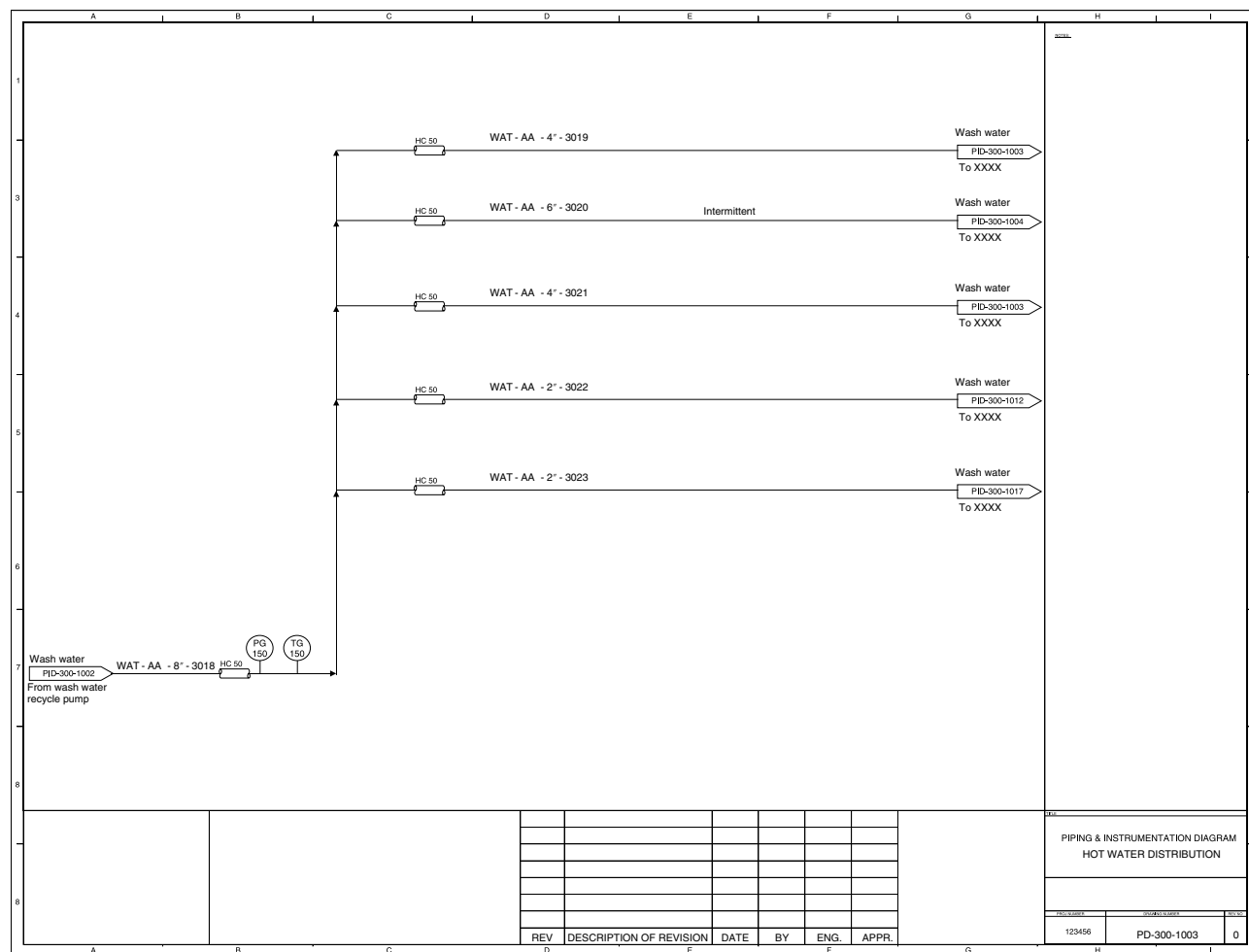


Figure 4.22 A Utility Distribution P&ID.

- Systems that are not directly related to the main process of a plant.
- Systems whose deletion from the main P&IDs does not hinder the understanding of the process (also cuts down on the crowded look) and eases readability.
- Systems that appear on several P&ID sheets and are exactly the same.

If a detail P&ID is referred to by several main P&IDs, the detail P&ID is named “typical detail P&ID”, but if a detail P&ID is referred to by one main P&ID, the detail P&ID is named “nontypical detail P&ID”. Examples of typical detail P&IDs are pump seal flush, sampling system P&ID, safety shower and eye-washer P&IDs, utility station P&ID, special control P&IDs (remotely operated valves, electric motors), piping detail P&ID, and HVAC equipment P&IDs. Examples of nontypical detail P&IDs are rotary machine lubrication and fire or gas detection and deluge system.

A referred detail of a P&ID can be written down near the system in the main P&ID, or the detail P&ID can simply be mentioned as one of several “reference” P&IDs on the main P&ID.

Some companies show the content of detail P&IDs in the guidelines and not on a P&ID. For example, a company may not like developing special control P&IDs and display relevant information in the I&C design guidelines. Piping detail P&IDs, which may have hook-up piping detail of sensors, steam traps, and instrument air manifolds, may not be provided in a P&ID and instead be mentioned in the Piping design guidelines.

Some of the types of detail P&IDs are explained below.

- HVAC drawings: These represent the HVAC system for industrial buildings. A sample HVAC P&ID is shown in Figure 4.25.
- Sampling system drawings: Sometimes the plant has many different sampling systems for the sampling of

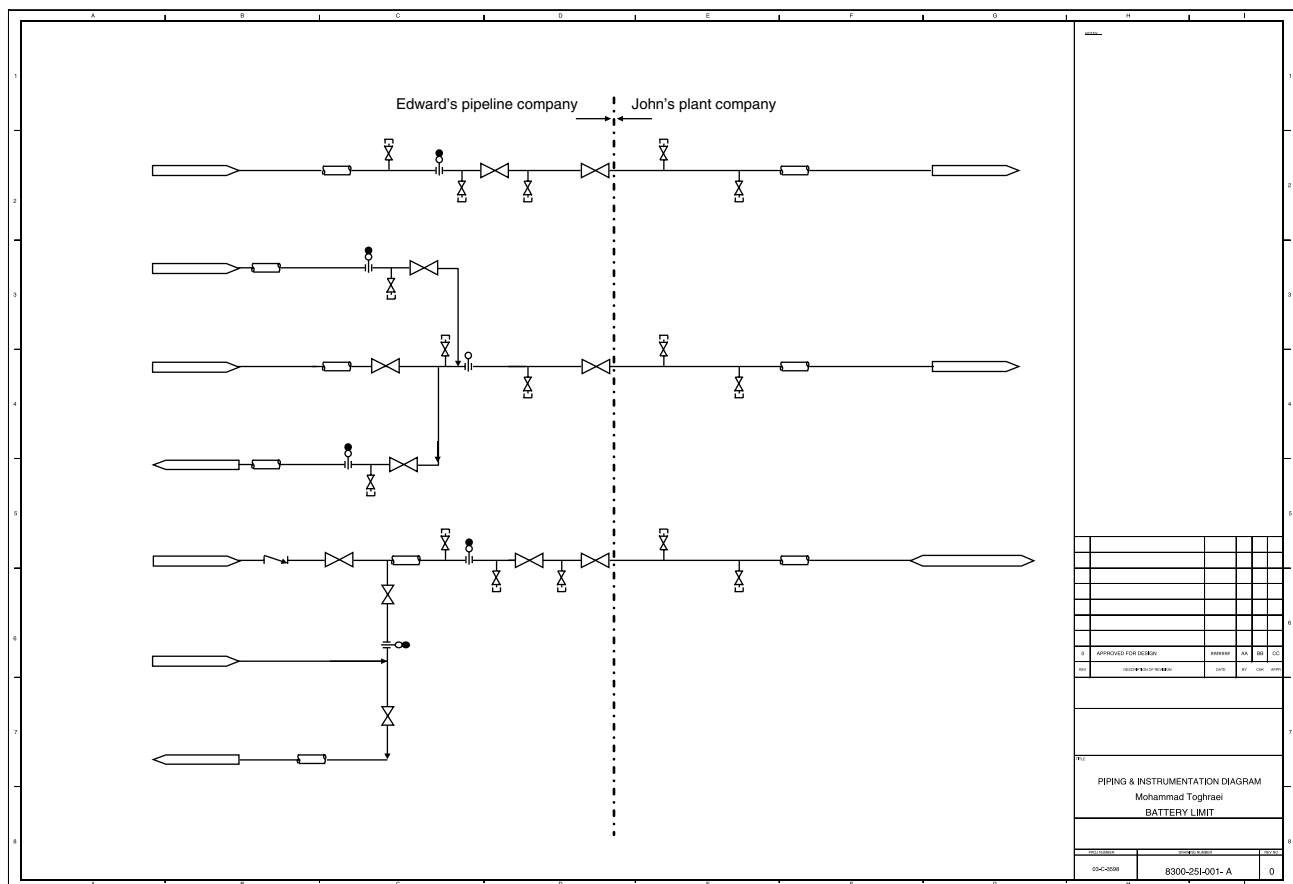


Figure 4.23 A Battery Limit P&ID.

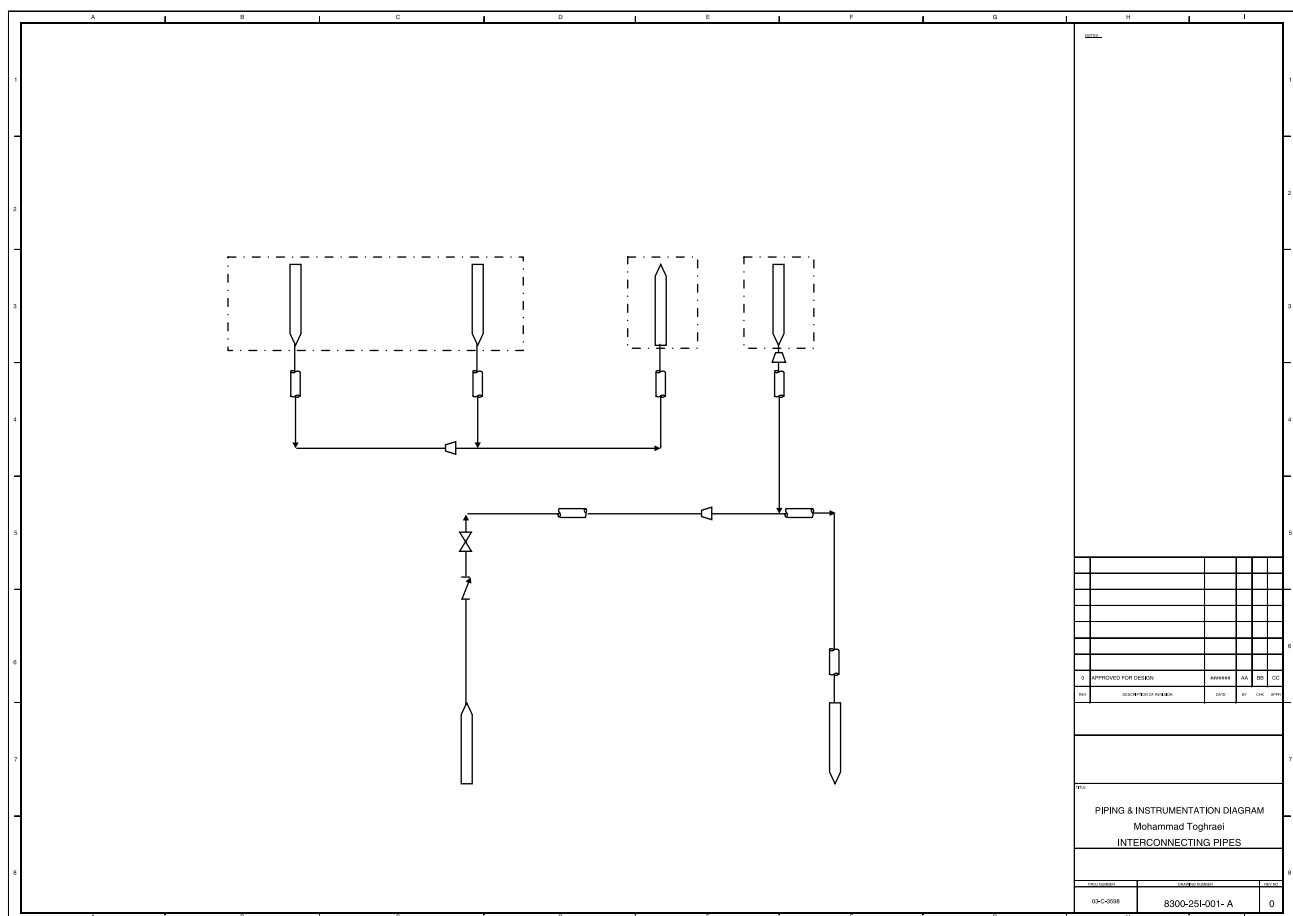


Figure 4.24 A Utility Distribution P&ID.

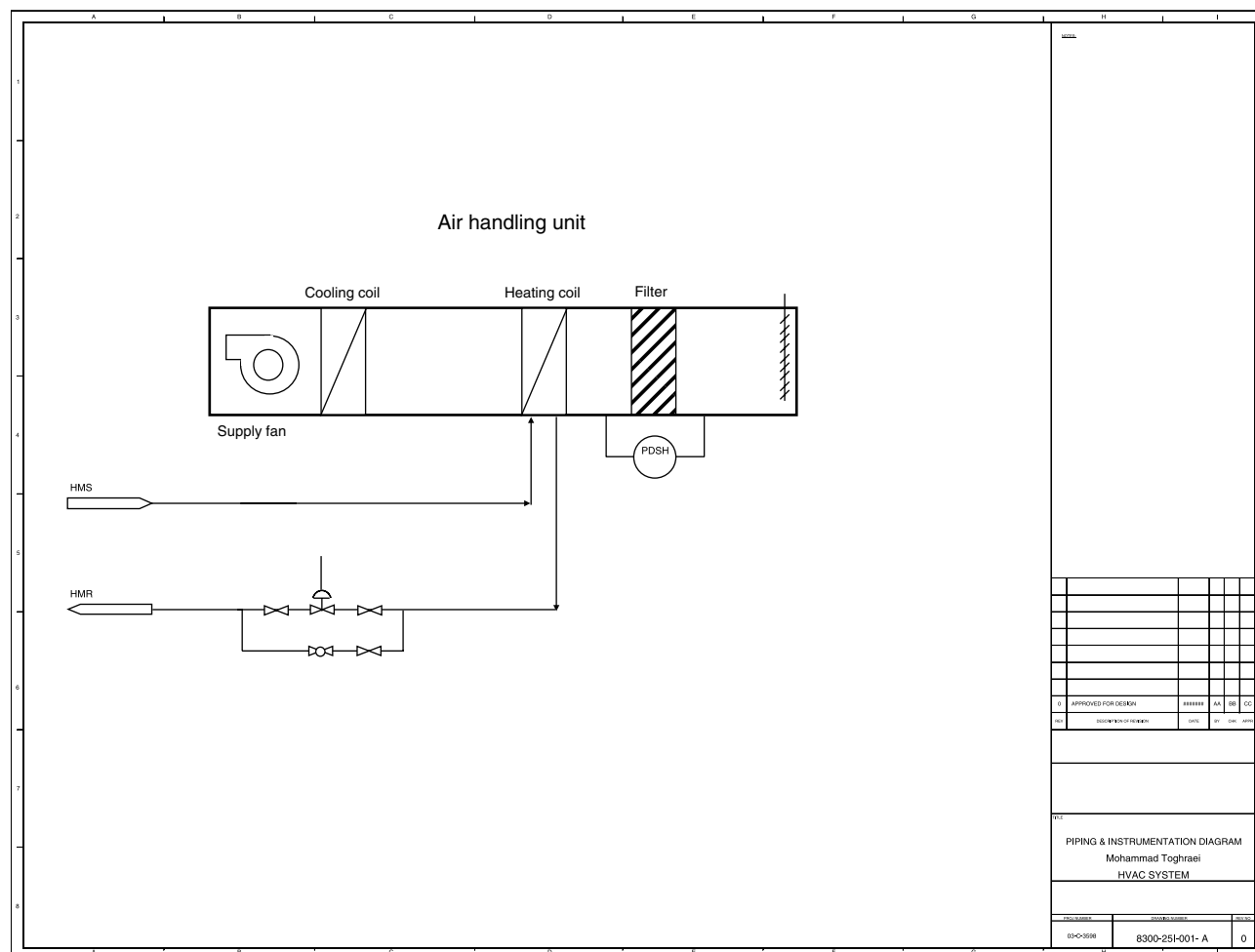


Figure 4.25 An HVAC P&ID.

gases, liquids, or even flowable solids. If this is the case, it may be all of the details of the sampling system are not shown on the main P&IDs and instead collected together and put at the end of the P&ID set; so what is seen on the main P&IDs may be only a “balloon” representing a typical sampling system. A sample P&ID sheet with a reference to a sampling system P&ID is shown in Figure 4.26, along with the sampling system P&ID (Figure 4.27).

- **Pump seal flush drawings:** A shaft generally penetrates a casing while it rotates. Examples of this are shafts for pumps, compressors, and mixers. The opening for the shaft should be loose enough to allow the shaft to rotate and tight enough to prevent any type of leaking. This problem is effectively solved by a mechanical seal. Mechanical seals are devices made up of two blocks of graphite sliding over each other. One block, which is shaped like a doughnut, is connected to the shaft, and the other block, in the form of a disk, is connected to the casing.

A mechanical seal needs to have a system for lubricating, flushing, and cooling. Such a system is called a seal flush system.

Sometimes companies decide not to show all the details of mechanical seals on the main P&ID to reduce crowding and because these items are not related to the main process of plant. In this case, the specific seal flush system used for a specific pump is mentioned on the main P&ID, and the reader is referred to the appropriate seal flush plan P&ID for the details. A sample P&ID sheet with a reference to a seal flush system sheet is shown in Figure 4.28, along with the seal flush P&ID in Figure 4.29.

4.5 A Set of P&IDs

A set of P&IDs may consist of different types of P&ID sheets. Figure 4.30 shows a typical order in a set of P&IDs. The legend P&IDs are logically the first P&IDs because they list the meaning of the symbols and acronyms. This

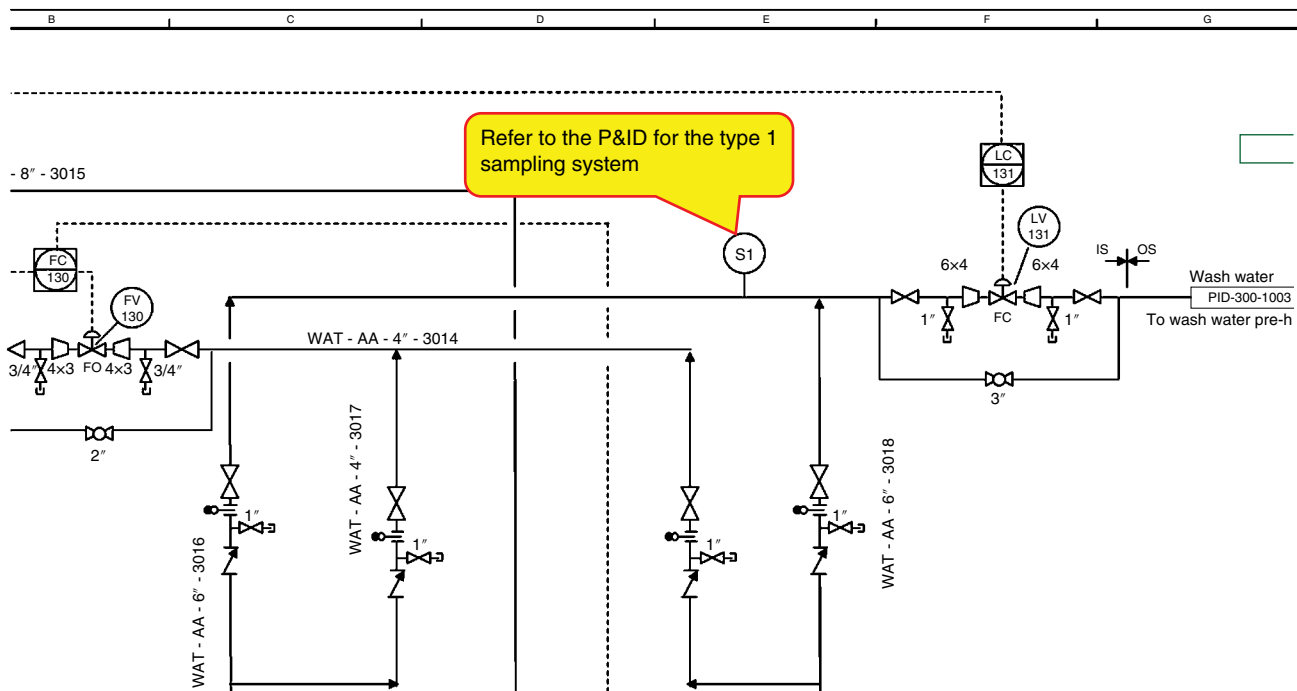


Figure 4.26 A P&ID sheet with a reference to a sampling system sheet.

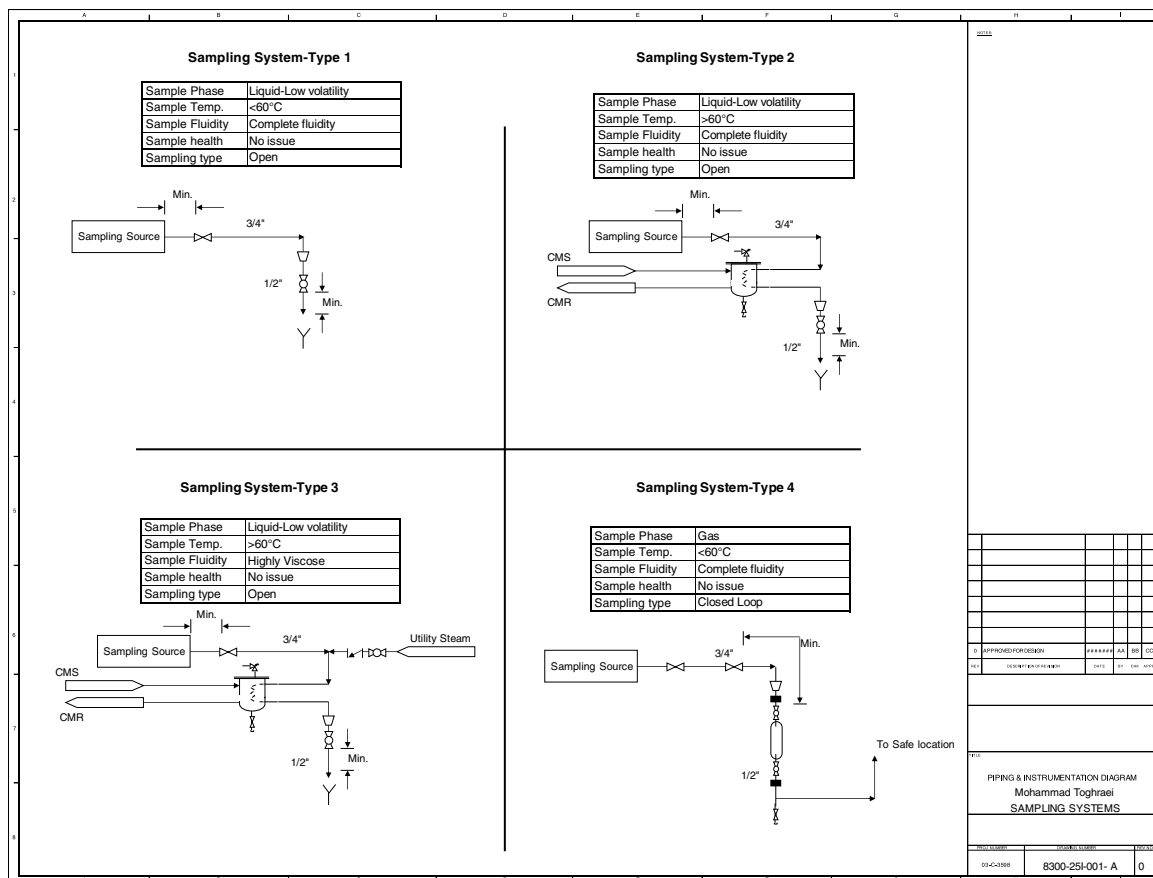


Figure 4.27 A sampling system P&ID.

Figure 4.28 A P&ID referring to a hypothetical seal flush P&ID.

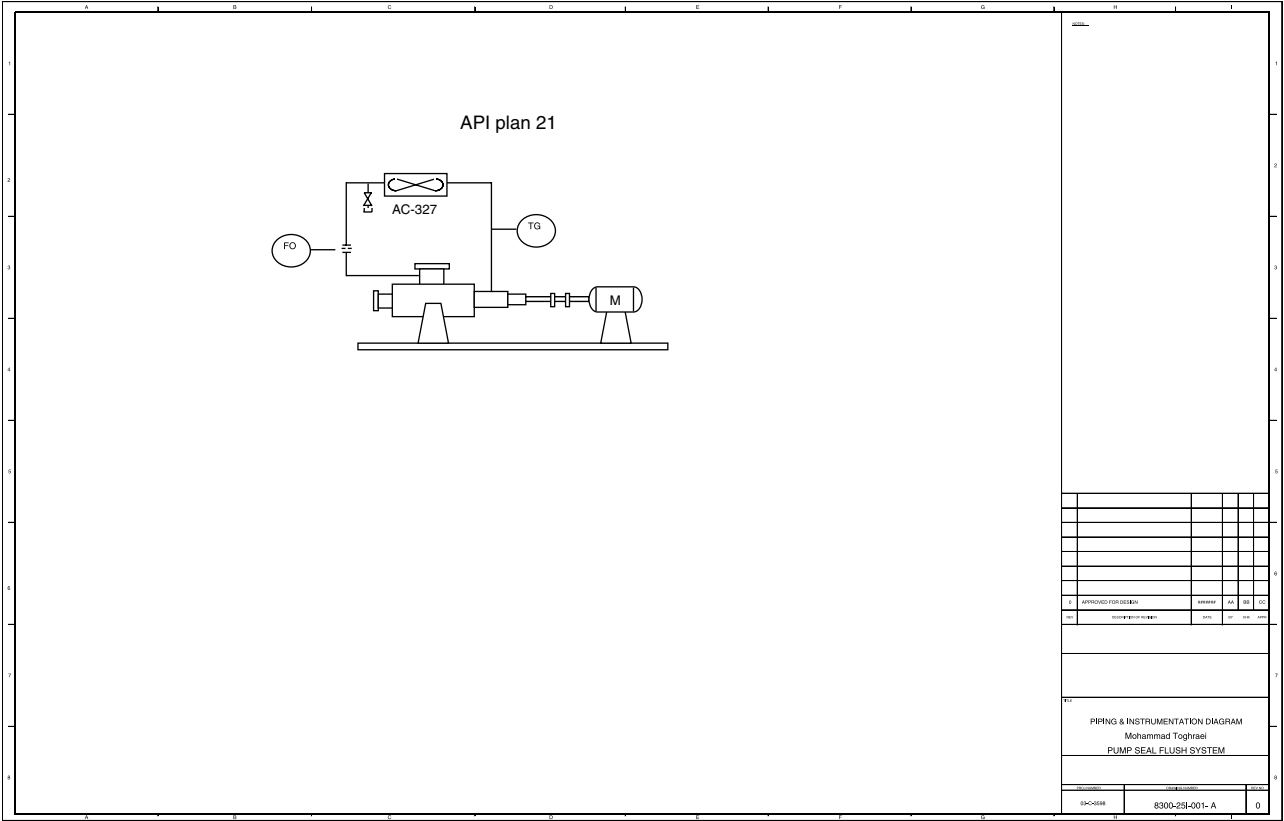
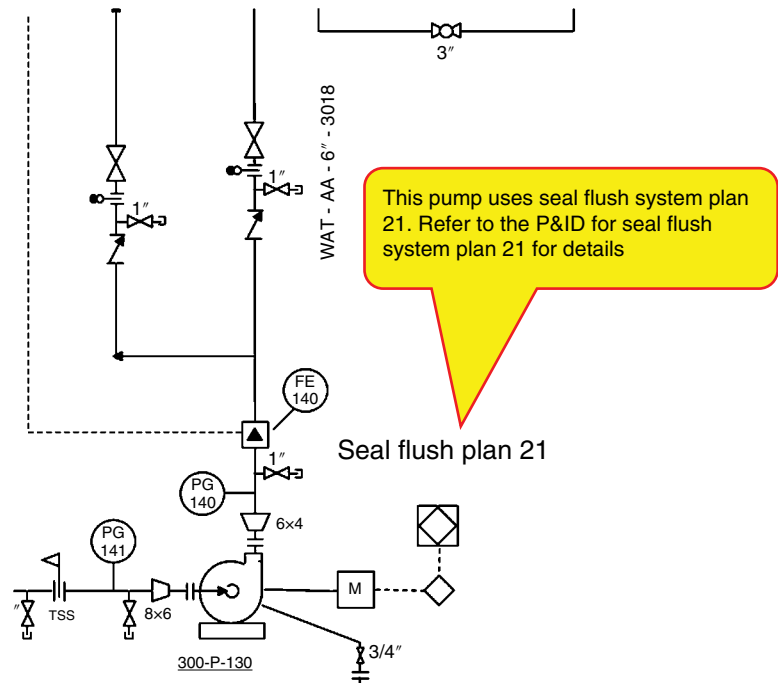


Figure 4.29 A Seal Flush P&ID.

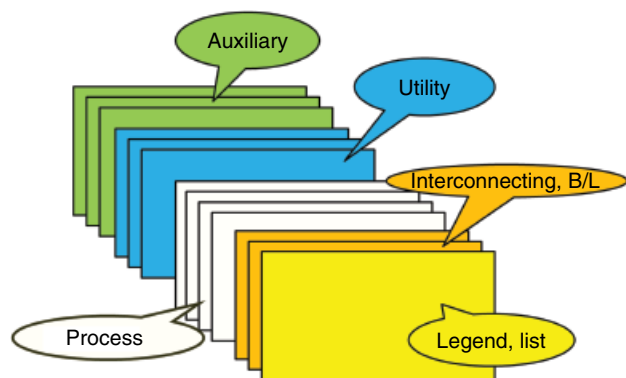


Figure 4.30 A given set of P&IDs.

is followed by the B/L P&IDs. The third and the main part of a set of P&IDs is the sheets that relate to the main process of the plant, or system P&IDs.

The process P&IDs show the route that the raw materials follow to be converted into product(s).

Utility and auxiliary P&IDs are the last groups of P&IDs.

In a design project, with this sequence, the network P&IDs are mainly dependent on the plot plan and the location of equipment should be finalized to be able to develop utility and interconnecting P&IDs.

All or majority of auxiliary P&IDs are created during the detailed engineering stage of projects, when the development of other P&IDs are near the end.

When designing a process plant, all the P&ID sheets of the plant should ideally be issued at once (simultaneously) as Issued for Construction (IFC). In the real world, however, such a thing may not be possible and figuring which P&IDs to issue first depends on the critical nature of the construction for the items on a specific P&ID sheet.

Generally speaking, pipe rack P&IDs should be issued first. P&IDs of large items are issued early, too, if they are not the vendor responsibility. The other high priority P&IDs are the ones for utility generation systems. Because the utility systems are usually the first systems that come into operation for commissioning, they should be constructed first. However if the utility generation systems are generic with low complexity, then they can be issued with lesser critically important P&IDs.

Such a priority in issuing P&IDs is for the IFC version only.

A P&ID set can be named based on not only its content but also its purpose. Each of the P&IDs mentioned thus far can be for a greenfield project or brownfield project. Brownfield projects can be upgrading or optimizing projects. In brownfield projects each of the discussed P&IDs can be converted to demolition P&ID and tie-in P&ID. In demolition P&IDs, the part of equipment that needs to be removed from the plant is specified somehow (e.g. hatched lines). In tie-in P&IDs, different tie-ins are added to show the pipes that need to be connected to a new item in the plant.

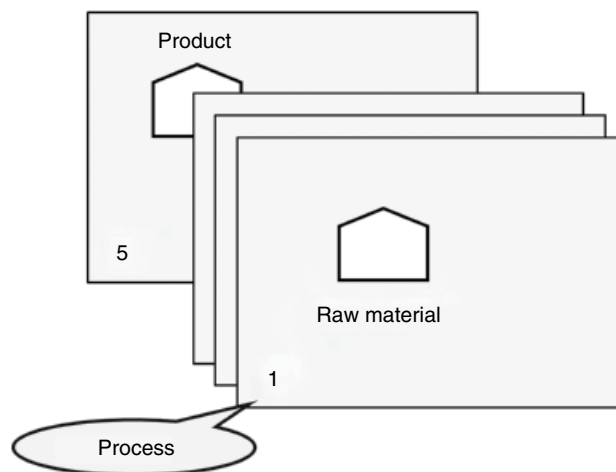


Figure 4.31 Process P&IDs within a complete set.

P-2600-1/2

Grundfos vertical inline
Centrifugal charge pump
CRN 64-2-2, 4"-150#RF,
339 USGPM • 125 ft head
c/w baldor electric motor,
15HP, 3450RPM, 254TC frame,
3PH/208-230/460V/60Hz,
Class I div II (TEFC)
Maximum discharge pressure:
190 ft head

Figure 4.32 Sample pump callout in a manufacturer P&ID.

4.6 P&IDs Prepared in Engineering Companies Compared to Manufacturing or Fabricating Companies

The P&IDs prepared by manufacturing or fabricating companies can be different than the P&IDs prepared by engineering companies. P&IDs created by engineering companies are prepared for the purpose of erection, installation, and start-up and should be kept in the plant for the life of the plant, whereas a P&ID made by manufacturing companies are prepared solely for construction.

The differences can be summarized as follows:

- 1) The P&ID set by manufacturing companies generally do not have auxiliary P&IDs. All the required details are shown on the main P&ID set. In many cases, vendors are not responsible for auxiliary systems.
- 2) The P&ID set by manufacturing companies tend to have more technical information. It is not strange to see the pressure range of a pressure gauge on a P&ID prepared by a manufacturing company. This is because manufacturing companies try to put as much as information on their P&IDs for other disciplines.

- 3) The P&ID set by manufacturing companies may have the brand name of items used on each P&ID sheet.
- 4) The P&ID set by manufacturing companies tends to have less design and operational notes. They may have less notes because of a closer relation with their design groups within their company. Also because the design groups work in one specific area, they are experts and already consider the details of design requirements, which are in the Notes block of other P&IDs.
- 5) P&IDs created by manufacturing companies may have more detail regarding instrument air.

The engineering company responsible for designing a process plant and developing the P&ID most likely does not have any item it built. The engineering company

[illegible]

Figure 4.34 Vendor-supplied loose items.

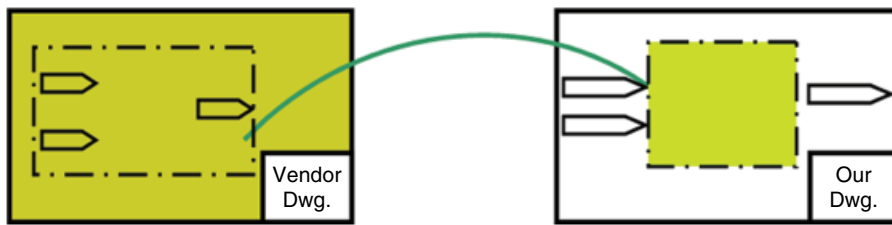


Figure 4.35 Integrating vendor P&IDs by implementing vendor P&IDs into master P&IDs.



Figure 4.36 Integrating vendor P&IDs by referencing.

A licensor can be defined as a specific type of vendor who registered his or her system as intellectual property and not all of the design can be disclosed to others, including clients and engineering companies. Vendor-supplied systems in P&IDs are generally shown using a dash-dot border around the system. All the connecting pipes (but not necessarily tubes) should show some sort of break (like a flange) at the vendor border line. This concept will be discussed in more detail in Chapter 6.

If there are items that should be provided by the vendor but cannot be surrounded by the vendor border, a predetermined sign (mentioned in the legend P&ID)

can show vendor responsibility for those items. The sign could be a simple star, or letter F (represents “furnished by vendor”). Figure 4.34 shows this concept.

These vendor-supplied items, which are outside of vendor border, generally are transported to the field as loose items. Engineering companies are not generally responsible for vendors’ or licensors’ designs, unless otherwise stated in the contract. Although the concept of showing vendor- or licensor-supplied items is clear, how companies incorporate these items on to the P&IDs vary.

The vendor P&IDs are sometimes implemented into the master P&IDs by redrawing them or “transplanting the snapshot” of their drawing (Figure 4.34). In other cases, the vendor border block remains as a blank block with only a reference sentence to the vendor P&IDs (Figure 4.36).

In all of these examples, except redrawing vendor P&IDs, it should be ensured that the vendor uses the approved symbols mentioned in the legend P&ID to avoid any confusion.

5

Principles of P&ID Development

This chapter attempts to answer the important question: whose benefit, and to what extent, should be considered in developing P&IDs? There is no universal set of rules for developing P&IDs. However, it can be said that the P&ID development rules come from the plant stakeholders.

5.1 Plant Stakeholders

The main goal of a plant is producing a product in the quantity and with the quality expected and designed for. However, there is more than that. A good plant design should first consider health, safety, and environmental codes and standards. The owner's wishes for the plant should also be noted, followed by the designer's and operator's requirements (Figure 5.1).

Jurisdictional health, safety, and environmental codes and regulations cannot be compromised. These rules are established to guarantee the safety and health of the society and protect the environment. It is important to realize the government regulations and codes are the "minimum" requirements, and some companies decide to go further to show they are good "corporate citizens". For example, the concept of sustainability is not always – and in its holistic approach – considered in government codes.

The owners of a plant, from an abstract point of view, are looking for a plant with the minimum required capital cost and the minimum expending/operating cost. They also want to have a plant that can be quickly built.

Designers like to work on projects in which the design bases are well established to make sure their design fits the purpose of the plant. If the design methods are unreliable, designers have two options. They may ask the client to do more research, tests, and piloting before starting a commercial plant, or designers can be asked to design in a way that addresses the different situations that may occur.

The operators are the firsthand end users of a plant and prefer one that can be easily operated and with reasonable flexibility and the highest level of safety. They may not always think about the capital costs or operating costs, but sometimes their instructions override the instruction of others, such as clients, engineers, and managers.

5.2 The Hierarchy of P&ID Development Rules

Each of the aforementioned parties initiates its requirements through different documents. Codes, regulations, guidelines, and standards are some of them. When two or more requirements exist, the designer (P&ID developer) should select one requirement based on the hierarchy of the rule makers shown in Figure 5.2.

Rule sets at the highest level are the ones set by jurisdictional bodies. These should be followed with *no* exceptions and are generally referred to as codes or regulations that cannot be overturned by any other rule maker.

Next are client guidelines, which may be applicable within the company or specific to a project. This is the highest level of rules that the designer should follow. The client rules sometimes can be waived if a proper "request for deviation" with adequate supporting documentations are submitted to the client and the client approves it.

The client may show less resistance to deviation if there is only company-wide (and not project-specific) guidelines.

Some clients are just small companies, which don't have much experience about the plant they want to build. Then there is no client guideline (neither project-wide nor company-wide) are available. In those cases, the designer refers to their own guidelines or the fourth level of rules or engineering company guidelines.

On the fourth level are the engineering company guidelines. Large engineering companies may have their own set of guidelines. They may get approval to use some

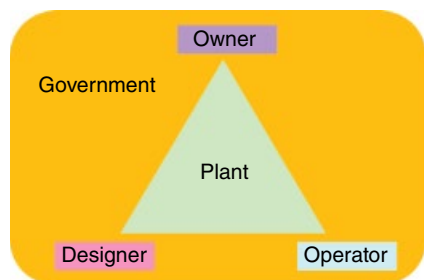


Figure 5.1 Plant stakeholders and their needs.

- Jurisdiction codes
- Client company-wide guidelines
- Client project guidelines
- Designer company guidelines
- General industry/engineering practices



Figure 5.2 Hierarchy of rule makers.

or all of their guidelines if there is no guideline for a specific topic in any other higher-level requirements.

At the last level are general industry practices. The most reliable source of general industry information is technical articles in engineering magazines, trade magazine, or peer-reviewed journals. The use of each source of guidelines is acceptable *if* the required information could not be found at the higher levels.

But before starting each project, the acceptable guideline sources should be communicated between the client and the engineering company, and these should be approved.

The most controversial phrases are *general industry practices* or *general engineering* because they can be different from everyone's point of view.

I had a meeting with a group of engineers and noticed that each person had an idea about the tank insulation based on general engineering practices. One believed that the tanks would only be insulated on their shells, if they were supposed to be insulated at all, and the another said that the shell and the roof of the tanks should be insulated.

General engineering practices can also be different for each fluid and each process. The nature of fluid affects the P&ID development. Fluids can be inert or fuming, toxic or flammable; can be liquid, gas, or flowable solids; can be perishable, precipitating, scaling, or fouling; or have single, dual, or even tri phases.

5.3 Plant Operations

Operations are the fundamental reason for having a process plant. Therefore, operations should be clarified at the beginning. The concept of plant operations is a

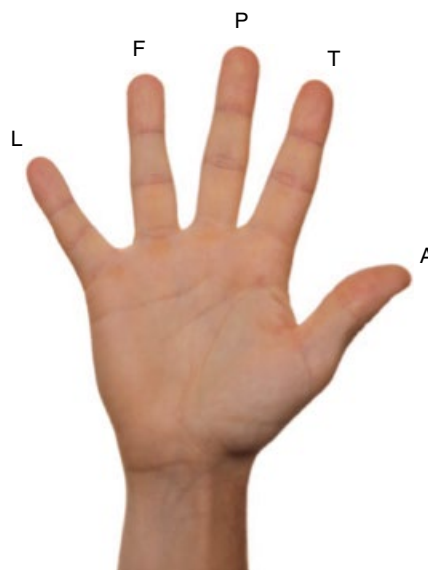


Figure 5.3 The most important process parameters.

common topic in P&ID development and in plant design. A person involved in plant, equipment, or instrument design or in P&ID development should be familiar this topic.

Plant operation relies on all the parts of the process parameters to function according to design. So, what are the process parameters and what are their different levels?

5.3.1 Process Parameters

In any process, there are a number of different parameters to consider. The five most important process parameters are as follows (Figure 5.3):

- **Level.** Level is defined only for nonflooded liquid containers and flowable solid containers. It cannot *usually* be defined for other process equipment, like pipes. Level can also be defined for open channels.
- **Flow rate.** Flow rate is a parameter that is almost exclusively defined for pipes. However, when talking about “flow of a piece of equipment,” we refer to the flow in the inlet or outlet of the equipment.
- **Pressure.** Pressure can be defined for any piece of process equipment, including vessels and pipes.
- **Temperature.** Once again, temperature can be defined and specified for any part of the process.
- **Analyte.** Analyte is a parameter that shows the composition. Unlike other parameters, composition is not one definite parameter. It consists of a number of items that can be measured with process analyzers. This might be the pH of water, the octane number of gasoline, the amount of hydrogen sulfide in gas, the brix in

a syrup, or the dissolved oxygen in water. Whatever it is, the composition parameter needs to be specified for each piece of equipment or part of the process. Here composition is defined as any parameter that directly or indirectly shows the composition of a stream.

In process plants, there may also be nonprocess parameters, but they are not as common. For example, the torque of a rotating scraper at the bottom of a sedimentation basin is a nonprocess parameter. Nonprocess parameters, if used, can be for monitoring or interlocking purposes. More detail is provided in Chapter 13.

5.3.2 Process Parameter Levels

To expand the preceding concepts more quantitatively, a seven-level milestone for each of the parameters can be defined as normal, low, low-low, high, and high-high plus two structure integrity levels (Figure 5.4).

Although these seven levels are not carved in the stone and some levels can be defined for any specific equipment and specific parameters, the seven milestones for parameters are common in process industries.

For the majority of engineers such a seven-level frame is obvious only for the liquid level in containers because it is generally placed on P&IDs for all containers. However, such a frame exists for other parameters, including pressure, temperature, analyte, and flow rate. The seven-level frame does not always exist officially and its “footprint” can be seen in an alarm table, shutdown key table, or other process documents.

These parameter levels are decided by the process engineer and are based on process engineering theories or client guidelines. Sometimes, however, specifying these operational parameters is not easy, and they need to be discussed with other stakeholders of the plant.

The normal level is the parameter constantly used during normal operation. It is actually the process optimum point at which the optimum process goal is achieved if

the parameter is kept on it. This level can be called the *normal operating point*.

The parameter can swing above and below the normal level up to the high level and down to the low level without the equipment or instrument losing any performance. That is the reason that sometimes high levels and low levels are called “normal maximum” and “normal minimum” levels.

At the high level, the system can still produce an acceptable product. This level is called the maximum process operating point, upset level, maximum allowable operating level, or maximum normal level.

Low level is similar to high level but on the lower side (lower value) of the parameter. The low level is called the minimum process operating point, upset level, minimum allowable operating level, or minimum normal level.

Generally, the process design of the system is done based on the high level parameter, and the operation is checked to make sure the system still works on the low level parameter.

When a parameter goes beyond high level or low level, the system is upset. This upset can be called a “mild upset” because the result is not only not achieving process goals but also losing the quality or quantity of the desired product(s). Therefore, the band of high level to high-high level and low level to low-low level is called a “mild upset” band.

When the parameter goes beyond high-high level or low-low level, the “severe upset” happens wherein the process goals have been lost and the current goal is to protect the equipment (hardware conservation) and the health and safety of personnel and neighborhood communities. A severe upset primarily threatens the integrity of the equipment. It can be classified further as maximum upset level and minimum upset level. A severe upset band refers to bands ranging from high-high level to high structural integrity level and from low-low level to low structural integrity level. When a parameter is in these bands, the performance of equipment will not be considered anymore.

The structural integrity level refers to the capability of a piece of equipment or instrument to hold without explosion or collapse of the body. It is important to know that parameters can still go higher than its structural integrity level without any issue as long as it is not on a continuous basis.

In these bands, no one is concerned about process goals because the process item or instrument may not even mechanically work anymore. The only vivid feature is the intact shape of the process item or instrument. In these bands, the process items and instruments are still standing and no major de-shaping happens.

The high and low structural integrity levels are determined by the equipment, pipe, or instrument manufacturer for design and manufacturing purposes. The process

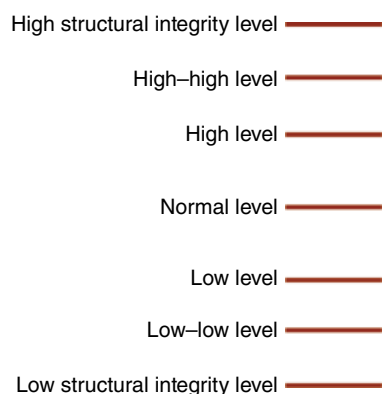


Figure 5.4 Seven levels for each parameter.

engineer will ask the mechanical engineer of the manufacturing company to design the equipment to withstand the high and low structural integrity levels. Therefore, these levels are named “design” values, too. However the word of “design” in this context only refers to the integrity of the equipment and not the operating features of the equipment. The high and low structural integrity levels are also known as “mechanical design parameters.”

When a parameter goes beyond the high or low structural integrity level, there is a potential of immediate danger as a result of an explosion or collapse of the process item or instrument.

Process parameters are arbitrarily split into four areas: normal operation, mild upset, severe upset, and immediate danger. These area (bands) are shown in Figure 5.5, and their features are outlined in Table 5.1.

By defining all process parameters for an item and propagating them for all parameter levels, a matrix will result that maps the operation of the item during lifetime of a plant. Table 5.2 shows an example of parameter definition matrix for a warm lime softener.

Process parameters on each level may have a specific name. They are not always named, for example, high-high level of pressure. These names are shown in Figure 5.6. A naming system is defined for each parameter, and these parameters are discussed more fully in the following discussion.

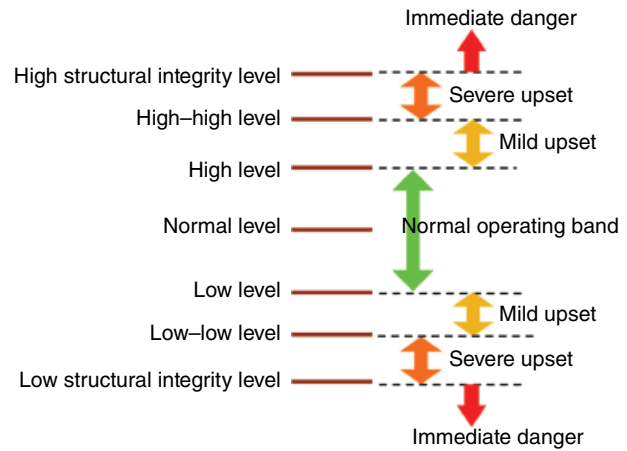


Figure 5.5 Main operation bands.

5.3.2.1 Pressure Levels

Pressure and temperature are the most important parameters for the structural integrity of all process items and instruments.

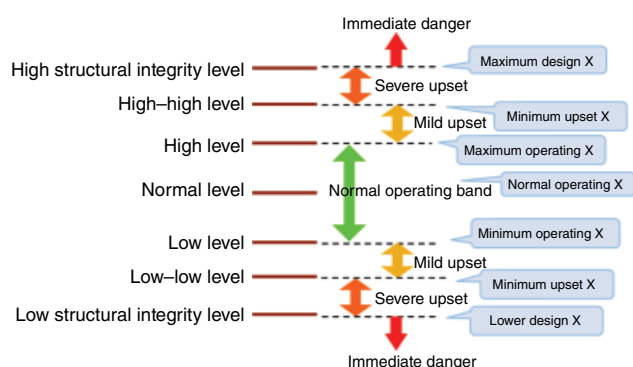
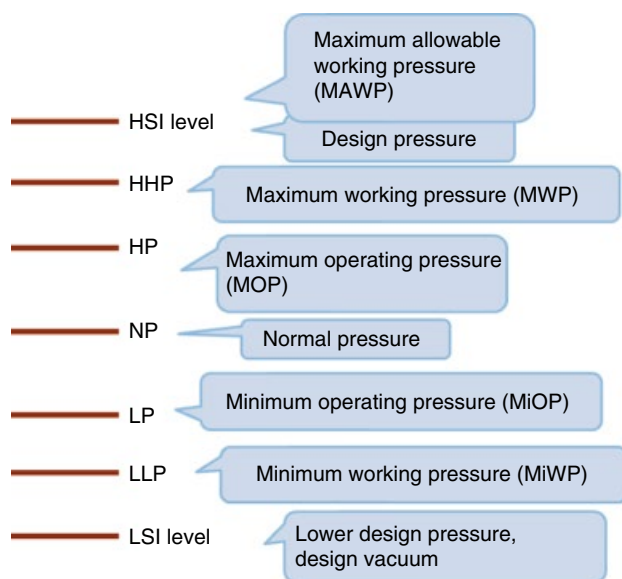
Design pressure is the value the process engineer needs from the mechanical engineer for the design of the structure of the equipment. However, the mechanical engineer of the fabricating company cannot necessarily follow the process engineer’s request because of many limitations, including the standard thickness of the

Table 5.1 The features of main operation bands.

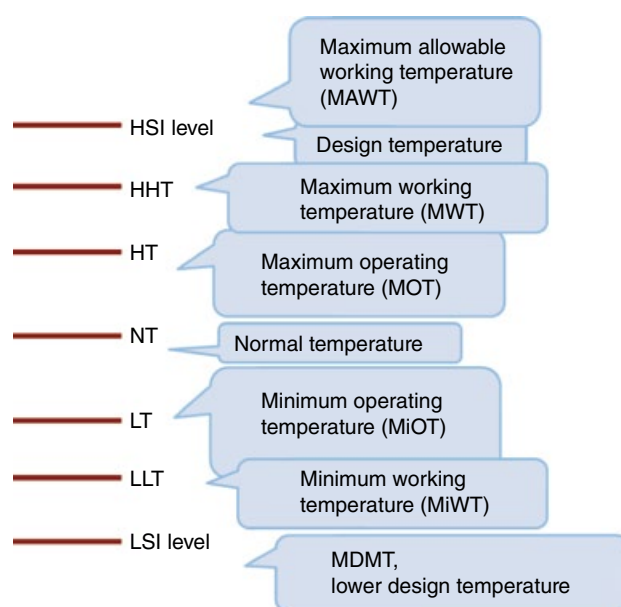
	Range	Process goal	Equipment functionality	Equipment integrity
Normal operation	High level to low level	Process goals are met	Equipment is fully functional and on its optimum window of operation	Equipment is intact
Mild upset	High-high level to High level or low-low level to low level	Process goals are not precisely met	Equipment is fully functional but not on its optimum window of operation	Equipment is intact
Severe upset	High structural integrity level to High-high level or Low structural integrity level to low-low level	Process goals are not precisely met; product could be off-specification or may not have product at all; hazardous material may be produced	Equipment is not fully functional; may not function at all	Equipment is intact
Immediate danger	Beyond structural integrity	Process goals are not precisely met; product could be off-specification or may not have product at all; hazardous material may be produced	Equipment is not fully functional; may not function at all	Equipment explosion or collapsing, release of gas, vapors, or liquids to environment

Table 5.2 Parameter matrix for a typical warm lime softener.

	Low structural integrity	Low-low	Low	Normal	High	High-high	High structural integrity
Temperature	−29 °C	60 °C	75 °C	80 °C	85 °C	100 °C	120 °C
Level	—	3000 mm	3000 mm	3000 mm	3100 mm	4000 mm	4300 mm
Pressure	−0.25 KPag	−0.15 KPag	−0.05 KPag	0.0 KPag	1.0 KPag	2.5 KPag	3.5 KPag
Flow (inlet)	—	180 m ³ h ^{−1}	195 m ³ h ^{−1}	200 m ³ h ^{−1}	205 m ³ h ^{−1}	220 m ³ h ^{−1}	Not applicable
Analyte (outlet)	—	No limit	No limit	20 mg l ^{−1} total hardness	25 mg l ^{−1} total hardness	30 mg l ^{−1} total hardness	Not applicable

**Figure 5.6** Names of parameters levels.**Figure 5.7** Pressure levels.

metallic sheet. The process engineer may design and fabricate the equipment based on a higher pressure called the “maximum allowable working pressure” (MAWP), which will be determined only after the fabrication of the equipment. During the design stage, only the design pressure is known (Figure 5.7).

**Figure 5.8** Temperature levels.

5.3.2.2 Temperature Levels

A temperature-level frame can be defined for everything such as equipment, containers, and pipes.

An interesting temperature level is called the Minimum Design Metal Temperature (MDMT), which is the lowest possible temperature that the body of a piece of equipment can experience.

MDMT, despite its name, can also be defined for non-metallic equipment. It relates to the metallurgy of the material the equipment is made of. If the temperature drops below this level, the structure of the equipment will be compromised. Sometimes the MDMT is set at minimum ambient temperature if there is no reason to use lower temperatures (Figure 5.8).

5.3.2.3 Liquid/Solid Levels

The liquid/solid-level frame is possibly the most well-known parameter-level frame because it is always mentioned on the P&IDs for tanks, silos, and nonflooded

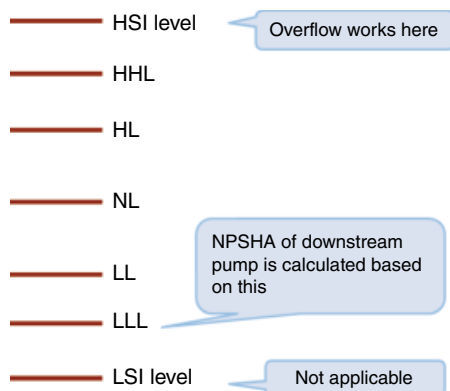


Figure 5.9 Liquid/solid levels.

vessels. This frame is not for gases. Figure 5.9 shows a liquid/solid-level frame with some examples for different levels.

Such a frame is more common for liquids and because of that there can be an extra L in the acronym for different levels that represent liquid. For solids, only a few levels can be used in the frame. It is mainly because of difficulties in measuring the accurate level of solids in silos, where there may only be a high-high level and low-low level.

5.3.2.4 Flow Levels

The level frame for flow is not as common as temperature and pressure. Flow is a parameter that is mostly defined for pipes. However, in cases where it is defined for equipment, it refers to the flow from the piping to the equipment. Figure 5.10 shows a flow-level frame with some examples for different flow levels.

Among the five-level frames of normal, high, high-high, low, and low-low levels, the last two are the most common. This is because equipment is generally more tolerant of high flow rates but may be more sensitive to low flows.

From a purely theoretical point of view, every piece of equipment can handle every flow rate, even huge ones!

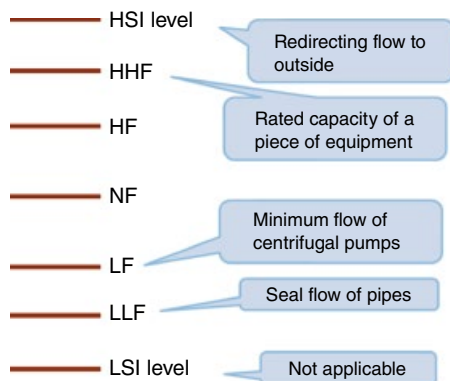


Figure 5.10 Flow levels.

With higher flow rates, there is only a higher pressure drop. If a piece of equipment cannot handle a specific huge flow rate, it is not because of the flow rate, but because the flow does not have enough pressure to overcome the pressure drop within the equipment.

For example, the low-low set point flow for pipes is the minimum flow that keeps the pipe full of fluid. Partial flow in a pipe not only is uneconomical but also can cause corrosion problems. One exception is gravity flow, in which a partial flow must be maintained. However, “seal flow” is so low that it is rare that anyone specifies low-low flow for pipes.

For some equipment, low or low-low flow rates are specified by the manufacturer. With a centrifugal pump, for example, a low flow rate is the flow rate generally specified by the manufacturer as the “minimum flow rate,” which means any flow rate less than that and pump will be instable because of internal flow circulation. Low-low flow rate could be defined as the flow rate that is not even able to fill the pump casing. Generally pump manufacturers do not bother to report this low-low flow rate because of its rarity. A low flow rate can also be a problem in fired heaters because the tubes can burn out.

5.3.2.5 Analyte Levels

A similar frame for analyte or composition can be defined. It is, however, not as common as other process parameters. Figure 5.11 shows such a frame for pH of a water stream.

5.3.3 Parameter Levels versus Control System

The control system will be discussed in more detail in Chapters 13–16. How can a control system work in a process plant? Let’s consider temperature as a process parameter in a warm lime softener as shown in Table 5.1.

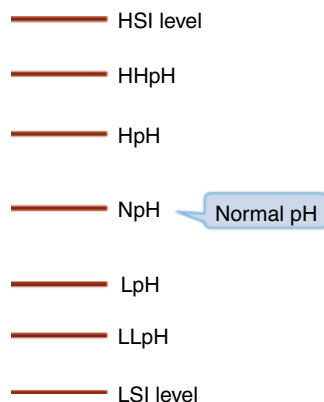


Figure 5.11 Analyte levels.

For the warm lime softener, we specified the normal temperature of 80 °C; it is *always* operated on this temperature. However, nothing is static, and the temperature will fluctuate with time. This may be because of a change in the ambient temperature, feed temperature, or feed composition. To maintain that temperature, a basic process control system (BPCS) needs to be installed for that piece of equipment. This system will attempt to regulate the temperature for the purpose of process smoothness, for example, by control valves.

That is why we specify a band around the desired temperature of 80 °C for the BPCS to operate in, say, from 75 to 85 °C. If a situation arises where the BPCS cannot control the temperature, and it rises or drops to a level outside this specified band, an alarm is activated. The alarm is meant to alert the operator to the danger and is an indication to take remedial action.

If, for some reason, the operator fails to respond in a timely manner or the action taken does not mitigate the issue, the Safety Instrumented System (SIS) will automatically come on and try to fix the situation. The SIS is not an adjusting, regulatory action like BPCS but rather a direct action, which may involve opening a valve, shutting down, or starting up a pump. It is a drastic action, intended to protect equipment from damage and to keep operators safe. In this example, the SIS kicks on when the temperature goes above 100 °C or below 60 °C.

If the SIS does not work to bring the process back under control, the temperature (the process parameter) moves to the final level. This is called a relief system, which is a purely mechanical system to protect the plant. Often, it may involve a safety valve that is activated to protect the plant, environment, operators, and other plant personnel. In this example, the safety valves can be set on 120 and –29 °C. Relief systems will be discussed in more detail in Chapter 12. This allocation of responsibilities among the BPCS, alarm system, and SIS is shown in Figure 5.12.

Now we can define the “bands” in regard to different parameter levels.

The band of high level to high-high level and low level to low-low level could trigger an alarm to warn the operator for the upset. When the parameter goes beyond these bands and passes into high-high (or low-low), a trip may be activated because it is in a severe upset band.

As you can see from Figure 5.13, the BPCS regulates and controls the process parameters between the high and low points. Therefore, the “playing court” for the BPCS is the band between the low and high levels. BPCS controls try to keep a parameter within low level and high level band. Once the parameter goes past these limits, this becomes a mild upset, and the alarm is activated.

If the operator is unable to rectify the situation in response to the alarm, the parameter may progress to the

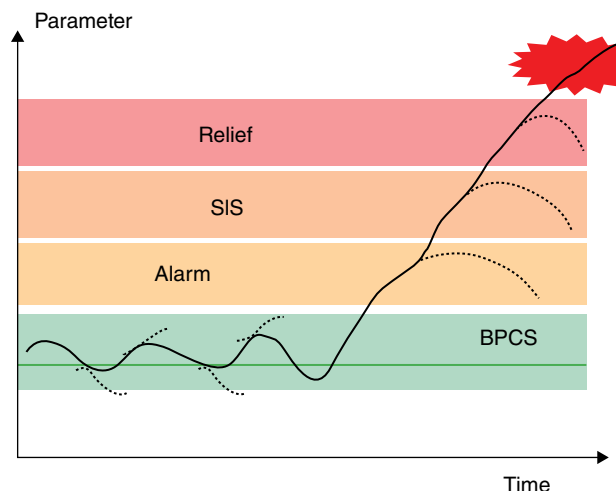


Figure 5.12 Process guard layers.

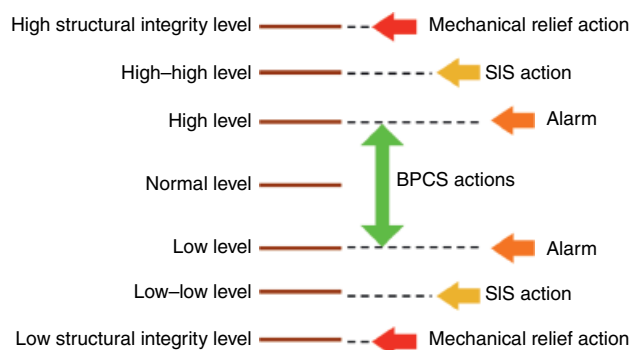


Figure 5.13 Action levels or bands for process guards.

high-high level or the low-low level. At these points, the interlock system, or SIS, will be activated.

Finally, if the SIS is not able to bring the process under control, a severe upset ensues. This is where the relief action is effected, usually through a pressure relief valve, to avoid an industrial accident. It is interesting to note that the BPCS works within a band, but an alarm system and the SIS are triggered at a certain point or points.

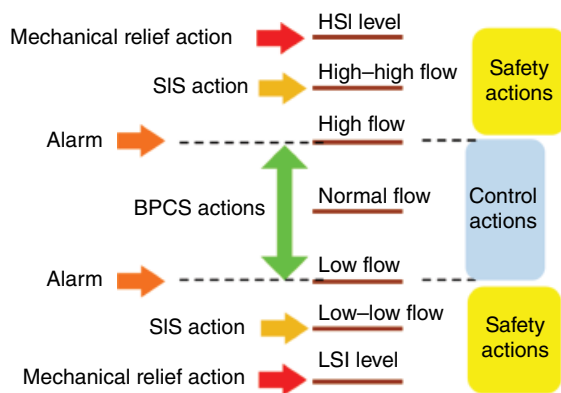
Table 5.3 summarizes these concepts for the example of temperature in warm lime softener. It is important to know that when the plant moves from BPCS to SIS, the actions of the SIS are not known as “control” functions. SIS works to prevent or mitigate the hazard. The next section discusses the relationship between different parameter levels and hazards.

5.3.4 Parameter Levels versus Safety

Generally speaking, when parameters are swinging within the band of high level to low level, the operation is safe. In a better terms, when parameters are in that band,

Table 5.3 Temperature levels.

Levels	Design consideration	Example
High structural integrity	Safety valve set point	120 °C
High-high	SIS action	100 °C
High	Alarm	85 °C
Normal	BPCS action band between low-low and high-high	80 °C
Low	Alarm	75 °C
Low-low	SIS action	60 °C
Low structural integrity	Safety valve set point	-29 °C

**Figure 5.14** Safety actions for a flow parameter.

there are enough provisions implanted in the system that we can consider the operation as a safe operation.

However, when the parameters exceed either level, some hazards start to be involved in the operation. Here

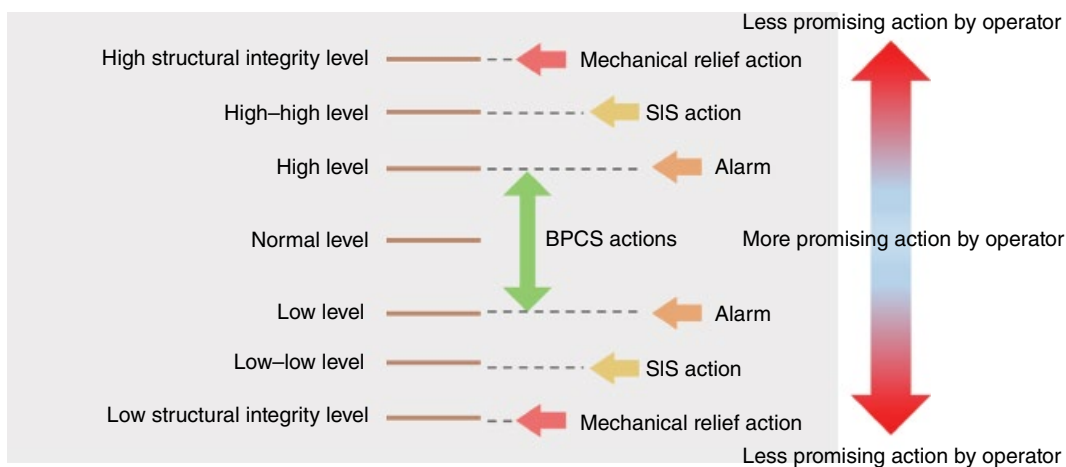
unsafe is the abridged version of the operation, which could be potentially harmful for the health and safety of people in and around the facility and also the environment. There should be alarms to warn the operator that an unsafe operation imminent. When the parameters reach high-high or low-low level, the interlock system (SIS) is activated to prevent harm to the health and safety of the personnel and save the environment from contamination. Figure 5.14 shows the level of parameter versus the level of hazard and control.

5.3.5 Parameter Levels versus Operator Role

Why do we need plant operators when we have all of these layers of control? As sophisticated as the control system may be, it is not intelligent (putting aside the concept of artificial intelligence for now). An operator with human intelligence to deal with an out-of-control situation is needed to run the equipment. It is this operator who should take remedial action to bring the process back under control and prevent the activation of the SIS. Remember, the SIS involves taking drastic and invasive action, which will interrupt the production process, with consequent loss in revenue for the company. However, it is vital to have the SIS layer built into the control system as a backup because an operator may not make the correct decision.

The sole purpose of an alarm is to alert an operator to a process parameter that is out of control and that it cannot be rectified by the BPCS. When an alarm is activated, the operator is expected to take action. It is essential that control system designers afford the operator every opportunity to respond; otherwise the alarm is pointless.

However, when a process parameter deviates from its normal operating band, the operator may be stressed and not make the best decisions (Figure 5.15).

**Figure 5.15** Operators' actions.

There are two main types of process plant operators: field operators and control room operators. When operators see an abnormal situation, they first try to mitigate it by stabilizing the system. If the system responds to the stabilization, the issue is resolved. Otherwise the second action is taken, or the operator reduces the capacity and slows the processes down. By slowing down, the operator tries to bring the issue to a manageable level. If none of this works, the operator starts to shut down the piece of equipment, the unit, or even the whole plant.

5.3.6 General Procedure of P&ID Development

The thought process for P&ID development has three main steps:

- 1) Addressing different phases of plant operation by adding more items.
- 2) Checking if the added items are not conflicting, and if any of added items can be merged together.
- 3) Dealing with common challenges in P&ID development.

Each will be discussed separately.

5.4 What Should a P&ID Address?

A P&ID should account for the full functionality of the plant in all stages of the plant life cycle, which can be outlined in four different phases: normal, nonroutine, inspection/maintenance, and the running without the item under maintenance.

For each process of the plant life phase, the three main elements of process plants – equipment, control system, and utility system – need to be designed properly.

- 1) Normal operation: In this phase all plant elements operate effectively and reliably. It means the functionality of equipment, containers, and piping are in their operating windows. In this step, the BPCS works to bring the item within normal conditions.
- 2) Nonroutine operation: In this phase, all plant elements operate but not “normally.” We can classify all these as nonroutine conditions. There may be low quantity of the product or a low-quality product (off-specifications product). In this stage, the SIS and safety relief valve may be triggered. This stage includes reduced capacity operation (system turned down), process upsets, start-up, shutdown, and any other phase of operation, which is *not* considered normal operation.

- 3) Maintenance and inspection: In this phase, the plant or some of its items undergo inspections and maintenance. Enough provisions for ease of inspection, rejuvenation, and maintenance should be provided. These provisions can include a wide range of things (i.e. isolation, draining, purging, steaming-out, or water flushing).
- 4) During maintenance on a piece of equipment, that piece of equipment is most likely nonoperational. The provisions should be provided to make sure the rest of plant is fully functional in the absence of that piece of equipment.

Each of these phases are expanded on in the following sections.

5.4.1 Normal Operation

For normal operation, each item on a P&ID needs to have enough capacity for its function. The main burden of this duty is on the equipment design and then control system, or more specifically, the BPCS. The equipment should be designed in a way that forces it to operate within a “window” of expected results, which is its best operating point. If this is not achievable through equipment design only (which is usually the case), a control system should be used on the equipment to bring the operation back into the operating window.

In a broader sense, a control system is supposed to bring the five main process parameters in the required range. These parameters are flow rate, pressure, temperature, level, and analyte.

Other than equipment design and control system (BPCS), heat conservation insulation also needs to be decided in this stage. All utility distribution and collection networks are designed for the purpose of this stage of plant life.

5.4.2 Nonroutine Operation

Nonroutine operations can be defined as any of the following situations but are not limited to them.

- Reduced capacity operation
- Reduced efficiency operation
- Start-up operation
- Shutdown operation
 - Planned shutdown
 - Emergency shutdown

In some plants, the provisions may also have a large effect on the design.

5.4.2.1 Reduced Capacity Operation

The capacity of the plant can occasionally be changed from the design capacity for a variety of reasons. It can be because of a shortage of raw material or u-transferred products in product storage tanks or having lost critical equipment or unit of plant as a result of failure.

Flexibility of operation in this context means the capability of a plant to operate in a wide range of flow rates without losing the quantity and quality of product(s).

One important point is flexibility; for example, reducing the capacity should be reached by a slow change to prevent any upset caused by fluctuation.

A plant is combination of equipment, utility networks, and control systems. To be able to design a plan with good flexibility, all of these elements needs to be high flexible. However, some of the elements need to be more flexible than others. Generally speaking, the control system (and specifically, control valve and sensors) should have the largest flexibility, and the equipment, the lowest flexibility. This larger flexibility for control items and utility network is because of the supporting role of the utility system and the controlling role of instruments in a plant (Figure 5.16).

There are two concepts that are used to quantify flexibility, one is the turndown ratio (TDR) and the other one is rangeability (R). The TDR is defined for equipment (including pipes and tanks) *and* also for utility systems, whereas rangeability is defined for instruments and more specifically for sensors and control valves.

When talking about flexibility parameters, there are two versions: provided and required. For example, the provided TDR of a piece of equipment could be 1.5/1 (by the equipment manufacturer), but the required TDR is 2:1 (by the plant owner). In such cases, the plant designer should use tricks to increase the flexibility of components. Table 5.4 shows the range of provided flexibility by different plant components.

In the following sections, the parameters representing the three elements of process plants – equipment, utility system, and control system – are defined. In the last section, different techniques to increase the flexibility of process plants are covered.

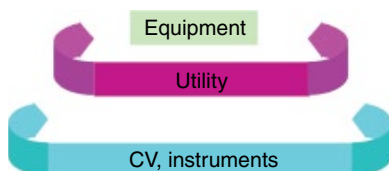


Figure 5.16 Required flexibility of different elements of a plant.

Table 5.4 Arbitrary required values of flexibility parameters.

	Low flexibility	Medium flexibility	High flexibility
Equipment (TDR)	<1.2:1–2/1	2:1–3:1	5:1–8:1
Instrument and control valves (rangeability)	≈ 4:1	10:1–30:1	20:1–100:1

5.4.2.1.1 Flexibility of Equipment and Utility: Turndown Ratio

A low flow/capacity operation of equipment or string of equipment may happen frequently in the lifetime of a plant. The reduced capacity operation could be intentional or accidental. The reduced capacity operation could be planned for off-loading the equipment for inspection, testing, and monitoring of the operation or even because of shutting down the downstream equipment. It also could be accidentally because of a drop in the feed flow rate.

The operator of a process plant likes to know how much the flow rate of the equipment (and in the bigger approach, capacity of the plant) can be decreased while still achieving the process goal and no off-specification product is generated. Therefore, the TDR can be defined as the ratio of low flow to normal flow.

$$\text{TDR} = \frac{Q_n}{Q_{Lo}}$$

where

Q_n is flow rate of system in normal level

Q_{Lo} is the flow rate in low level

The numerical value of TDR can be reported as ratio $\left(\frac{2}{1}\right)$.

It is important to consider the denominator term is the flow rate in the low level and not low-low level. Because the flow rate in the low level is generally considered the minimum level of flow at which the process goal is still reached. Figure 5.17 summarizes the concept of different flow rate levels and two terms of flexibility.

The TDR is sometimes mentioned in the client's documents for design purposes, but usually operators are looking for at least a TDR of 2:1 for a plant. The required TDR could be as high as 3:1 or 4:1 though.

When a centrifugal pump with the capacity of $100 \text{ m}^3 \text{ h}^{-1}$ has a minimum flow of $30 \text{ m}^3 \text{ h}^{-1}$, it means that the centrifugal pump has a TDR of 3:1 (without minimum line backflow system).

The TDR of a reciprocating pump can theoretically be defined as infinite because it can work in a wide range of flows. However, in practice, the pumps cannot handle

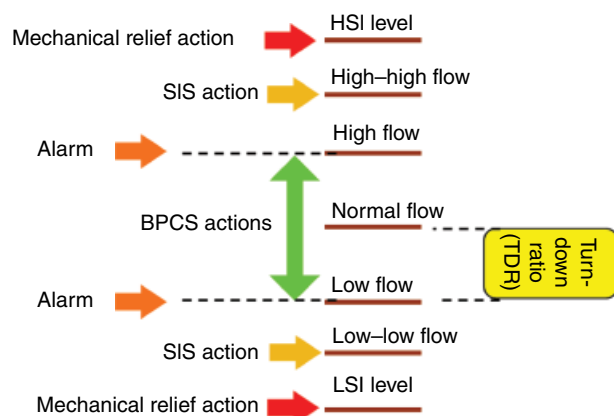


Figure 5.17 Concept of turndown ratio.

the flow rates that fail to fill the cylinder of the pump in one stroke. Partial filling of the cylinder may cause some damage to mechanical components of the pump in the long term. This minimum flow is a function of cylinder volume and stroke speed of a specific pump.

The TDR of a pipe is a bit more complicated. Here the minimum flow should be defined as the minimum flow that does not fall into laminar flow, or the minimum flow that keeps a check valve open. For liquid flows in pipes, the minimum flow can be interpreted as the minimum flow that makes the pipe seal (no partial flow pipe) because a flow smaller than that will freeze in an outdoor pipe. If the flow contains suspended solids, the minimum flow should prevent the sedimentation of suspended solids.

Table 5.5 is a table of typical TDR of some equipment. Table 5.6 shows the large TDR of storage containers. The high TDR of these items also explains why containers are used for surge dampening in plant-wide control practices.

In some cases, deciding on the required TDR needs additional consideration. One example is a chemical injection package. TDR is important for chemical injection packages to ensure that there is no time that chemical overdoses or underdoses happen, if both of them are intolerable to the process.

It is popular to expect to see a chemical injection package provide a TDR of about 100:1 or lower; possibly 10:1 can be provided by stroke adjustment and another 10:1 through VFD. But why is such large TDR necessary if the host flow experiences only 2:1? This huge TDR is generally because of the uncertainty in required chemical dosage and abundance of chemical producers. If the dosage is fairly firm and the chemical is a nonproprietary type, the TDR can be decreased to lower the cost of the chemical injection system.

Table 5.5 Turndown ratio of some selected equipment.

Item	Turndown ratio
Pipe	Large but depends on the definition of maximum and minimum flow.
Storage containers (tanks or vessels)	Very large (maximum is total volume of the container, but minimum could be dictated by downstream item. For example a centrifugal pump dictates a minimum volume to provide required net positive suction head [NPSH]).
Centrifugal pump	Typically 3:1–5:1
PD pump	Theoretically Infinite
Heat exchanger	Small, depends on the type (e.g. less than 1.5:1)
Burner	Depends on the type between 2:1 to 8:1

Table 5.6 Utility surge container to provide turndown ratio.

Utility	Surge container
Instrument air (IA)	Air receiver
Utility water (UW)	Water tank
Utility steam (US)	Steam drum in conventional boilers (in steam generators cannot be stored; the system design should be in a way to “float” US with other streams.
Utility air (UA)	No dedicated container can “float” with IA
Cooling water (CW)	Cooling tower basin
Cooling or heating glycol	Expansion drum

5.4.2.1.2 Flexibility of Utility Networks

The flexibility of utility networks is also defined by TDR. As mentioned, when a TDR of, for example, 2/1 in a plant is requested, the TDR of utility network should be higher.

As the utility system needs a large TDR, it generally needs a container in the utility production area to absorb the fluctuations caused by the utility usage change in process areas. Table 5.6 shows the name of these surge containers in different utility systems.

The utility network by itself experiences different levels of turndown, and consequently it needs different TDR. The main header can need the minimum TDR, whereas the subheads may need a higher or lower ratio (Figure 5.18).

Achieving a high TDR for the utility network and instruments is not difficult. The utility network is mainly

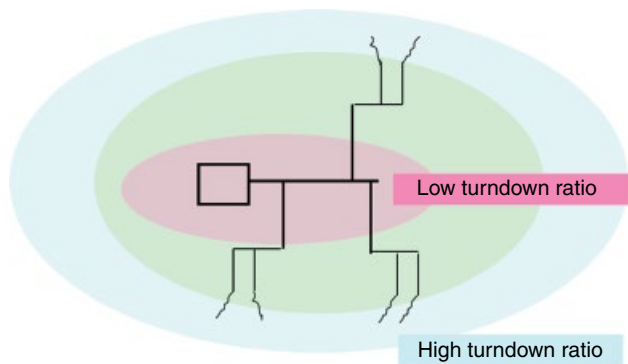


Figure 5.18 Map of turndown ratio for a typical utility network.

Table 5.7 Turndown ratio of selected instruments.

Item	Turndown ratio
Flow meter	Depends on type and between 3:1 to 50:1 and more.
Control valve	Depends on type and the characteristics, generally 50:1, and less than 100:1

composed of pipe circuits, which inherently have long TDR, and instruments (control valves and sensors) have intrinsically large TDR generally better than 20:1.

5.4.2.1.3 Flexibility of Control Valves and Sensors: Rangeability

Instruments have a higher level of responsibility than the equipment and the utility network because their duty is not limited to normal operation or a band of low to high levels. They have to be operational in a wider band of low-low to high-high. Therefore, rangeability can be defined as:

$$R = \frac{Q_{Hi\ Hi}}{Q_{Lo\ Lo}}$$

where $Q_{Hi\ Hi}$ is flow rate of an instrument or control valve in high-high level and $Q_{Lo\ Lo}$ is the flow rate in low-low level.

Table 5.7 is a table of typical rangeability of some instruments.

5.4.2.1.4 Providing Required Flexibility

There are three main ways that specific flexibility for plant items can be given: by using items with inherently high flexibility, placing similar items with a smaller capacity in parallel, and providing a recirculation route.

One example of using recirculation to provide enough TDR is the minimum flow line for centrifugal pumps. Using multiple pieces of equipment in a parallel series to

provide enough TDR is popular in designing sand filters in the water treatment industry.

- 1) Using equipment that has inherently higher flexibility. There are some process elements that have inherently higher flexibility such as tanks and pipes.

It is not always easy to recognize if a piece of equipment has inherently high or low TDR; however, there are some guidelines that can be used to determine this:

- Small volume equipment have a slimmer TDR than larger volume equipment
- Equipment with internal baffles (compartmented) have lower TDR.

An example of this is some gravity separators such as oil–water separators.

- Equipment in gas services may show higher TDR than the ones in liquid services.
- Equipment with internal weir (especially fixed ones) may have very low TDR.
- Equipment that uses some properties of inflow for its functioning may have lower TDR. Examples are a cyclone and hydrocyclone. In cyclones, the energy of the flow is used to generate centrifugal force; less flow causes less centrifugal force and that may cause an ineffective system. Another example is conventional burners; the lower fuel to a burner may cause short flame or even back fire in the burner.

The utility network should have a large TDR, and they are mainly comprised of pipes in different sizes. If a control valve is necessary in the network, sometimes it may need to have parallel control valves with split control because of the large required TDR.

- Equipment containing loose porous media may show lower TDR in liquid services. And the TDR may be lower when the porous media is built with larger particles.

Examples of these types of equipment are sand filtration, catalyst contactors, and the like.

Instruments generally have an inherently high rangeability, which satisfies the plant requirements most of the time.

- 2) Using parallel equipment.

Instead of using equipment with the capacity of $100\text{ m}^3\text{ h}^{-1}$, use an arrangement of two parallel equipment each with the capacity of $50\text{ m}^3\text{ h}^{-1}$. By doing this, a TDR of *at least* 2:1 can be used (*at least* because the equipment may have inherent RDR capability that may be added to the provided 2:1 TDR).

By using three, $33\text{ m}^3\text{ h}^{-1}$ pieces of equipment in a parallel arrangement, a better TDR of 3:1 can be achieved. This technique has other benefits; the parallel arrangement provides higher availability for the system because losing two or three parallel equipment is less probable than losing just one. Instruments generally

have a high and suitable flexibility (rangeability). But if their rangeability is not enough, generally using them in parallel is the solution. For example, using two control valves in parallel in a single control loop of “split range” is example of this technique. However, there are some disadvantages involved, including the increase in the capital cost and operating cost.

3) Providing recirculation pipe.

This method is exclusively used for noninstrument items. Implementing a recirculation pipe from an equipment outlet to its inlet is a popular method to increase the TDR of a system. The recirculation pipe may need some control system placed on it to prevent full flow backward. An example is using a minimum flow line for a centrifugal pump. A centrifugal pump with the capacity of $100\text{ m}^3\text{ h}^{-1}$ and a minimum flow of $30\text{ m}^3\text{ h}^{-1}$ (i.e. a TDR of 1:3) can be equipped with a minimum flow line with an appropriate control system to increase its TDR (Figure 5.23). If the minimum flow line and the control system are designed to handle a maximum $30\text{ m}^3\text{ h}^{-1}$, it means that the TDR of the pump is theoretically increased to infinite by zeroing the minimum flow. However, this method cannot be applied to all types of equipment. For example, this is not a good technique to increase the TDR of a furnace or fired heater;

recirculation of fluid around a furnace may increase the furnace coil temperature, causing it to burn out.

5.4.3 Reduced Efficiency Operation

Reduced efficiency of a piece of equipment or unit may generate low-quality or off-specification products.

If the quality of product is still acceptable, nothing can be done. Also nothing can be done if there is already a higher-than-needed quality product mixed the low-quality product because this still creates an acceptable product. In this case, though, a product tank (or any other product bulk system) is needed to allow the mixing of high-quality with low-quality products.

The low-quality products can be routed back to an upstream point to reprocess them to get acceptable products. They can also be redirected to an off-specification tank and then mixed gradually with overly high-quality product. The low-quality products can be routed to an off-specification tank to rerun the batch and turn it into high-quality product, returning it back to the system when complete. Low-quality products can also be sold to a third party for further processing to attain the high-quality product or other marketable product. Table 5.8 outlines these techniques.

Table 5.8 Options for dealing with a low-efficiency operation.

Option	Schematic
1 Do nothing!	No schematic; no any specific addition to P&ID
2 Recirculating the off-specification product to upstream unit	
3 Redirecting the off-specification product to an off-specification tank and then returned gradually	
4 Redirecting the off-specification product to an off-specification tank, retreat it (by batches) and then return it	
5 Redirecting the off-specification product to an off-specification tank and then sell it to a third party for retreating it or selling as is	

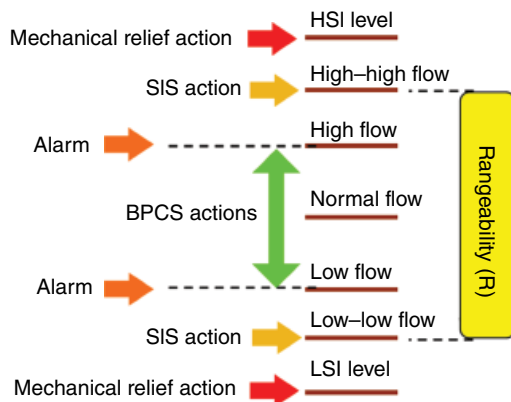


Figure 5.19 Concept of rangeability.

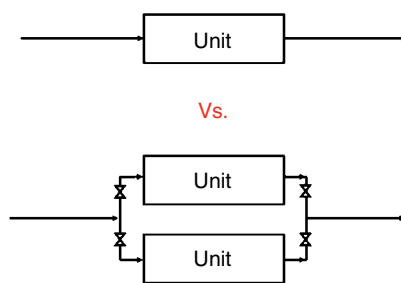


Figure 5.20 Using parallel equipment to provide a required turndown ratio.

These actions can be done manually after an operator or an automatic control system checks the quality of the product. Any of the aforementioned solutions can be used. However, the more important point is that the cases of low-efficiency operation should be considered for each unit that creates a physical or chemical change on the process stream.

5.4.4 Start-Up Operations

There are at least two types of start-up operations: the first start-up of a plant after its construction, which is called *commissioning*, and the start-ups after each shutdown.

The start-ups after shutdowns can be in two different types: start-up after a *planned* shutdown and start-up after *emergency* shutdowns. A start-up after an emergency shutdown may have more steps than a start-up after a planned shutdown.

Start-up operations can be assumed to be a severe capacity reduction case. In this situation, not all the instruments will work properly because process parameters during the start-up are not necessarily within their range. However, the operating personnel who are

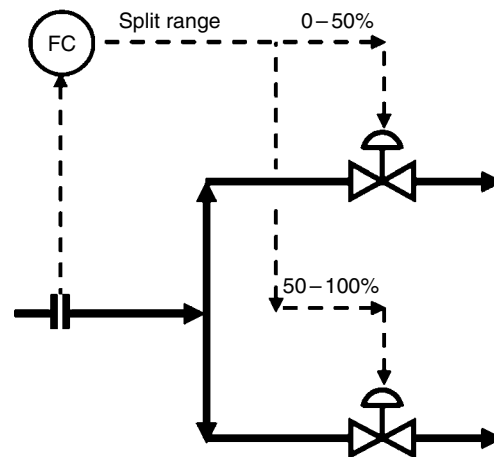


Figure 5.21 A control valve arrangement in wide rangeability requirement.

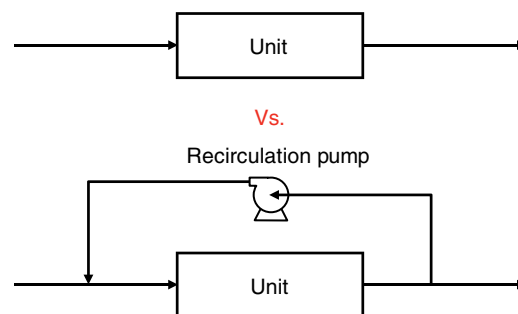


Figure 5.22 Recirculating for increasing TDR.

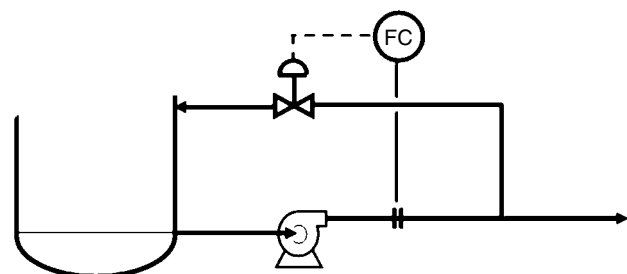


Figure 5.23 A centrifugal pump minimum flow recirculation to increase TDR.

attending the plant are the resources available and expected to compensate for the lack of instrument operability. They will be present as a larger group during normal operation.

The commissioning operation is a specific case of a start-up. In addition to all issues related to a start-up, a commissioning has many other problems related with poor construction and installation. During the plant construction, the electric motors may be connected to

rotate in reverse, the bolts may not be fastened enough, thus leaking, and there can be a junk left inside pipes.

A good P&ID should take care of these issues, too. For example, a temporary suction strainer (TSS) can be installed on the suction side of centrifugal pumps to protect the impeller from incoming debris during commissioning. The strainers should be removed from the pipe later.

A general procedure for starting up of a process equipment is as follows:

- Venting or draining the system: A system ready to be started can be filled with process fluid (gas or liquid), it can be empty, or it can be filled with air. In any case, the system should be empty at the start.
- Inerting: This step can be skipped in some cases. The goal is to replace any dangerous atmosphere with an inert and readily removable atmosphere. This is done by introducing an inert gas to the piece of equipment.
- Warming up: This step is needed if the operation temperature of the piece of equipment is different than the ambient temperature. To do this, a small amount of process fluid is allowed to into the piece of equipment.
- Partial loading and recirculation: Start-up operations can be done by recirculation. Figure 5.24 shows the basics of this procedure.

If the system is not reversible, the start-up operation can be more complicated and case specific.

If the start-up procedure goes through the recirculation (which is not a rare case), the highest attempt should be made to avoid using a long pipe for the purpose of start-up. It is not a sound decision to spend lots of money for the pipe that is supposed to be used only during the start-up. As much as possible, the existing pipe arrangement should be used for the purpose of start-up recirculation, especially when a high-bore pipe is needed.

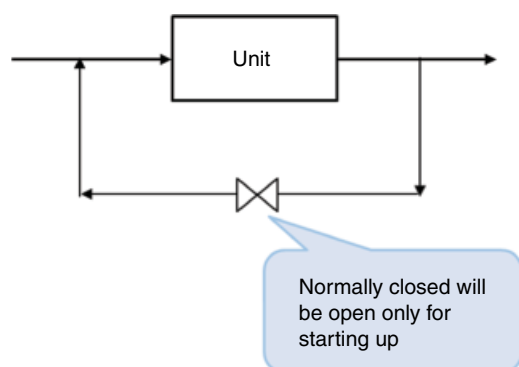


Figure 5.24 General procedure for starting up a unit.

The tendency of using the piping arrangement, which was implemented for normal operation, for the purpose of start-up recirculation is so strong that some process engineers forget to think about the start-up operation during the development of the P&ID and assume that they will find a way to accommodate start-up somehow.

5.4.5 Shutdown

There are two types of plant shutdown: planned and emergency. The goal of a shutdown is stopping the production of a plant or a unit. However, a planned shutdown happens based on a prior decision and preparations, whereas an emergency shutdown occurs because of circumstances beyond anyone's control.

The aims of a planned shutdown are upholding operators' safety, protecting the hardware asset, and minimizing or preventing product loss. The restart-up after a planned shutdown is easy.

An emergency shutdown has the same aims; however, minimizing or preventing product loss may not be achievable. One aim of the interlock system is to complete an emergency shutdown in a way similar to a planned shutdown. During an emergency shutdown, the main driver of a process plant can be interlock system (SIS) because in such situations, the operator's actions are not reliable.

The reasons for an emergency shutdown are diverse. However, because every process plant operates based on cooperation among the elements of equipment, utility system, and instrumentation and control system, an emergency shutdown can be classified based on the failure of each those elements. The error by operators or other plant personnel is the fourth source of emergency shutdown.

Failure or rupture of a tank may lead to an emergency shutdown. Fire in a limited area of a plant may also lead to an emergency shutdown because of failure in equipment, utility network, or instrumentation and control system.

One very common emergency shutdown is because of utility failures. For example, an electricity blackout may cause an emergency shutdown. In many cases, an emergency shutdown leaves stopped equipment with fluid contained in them. Such a situation is considered potentially problematic. The main issue about equipment with fluid in it is setting the fluid inside of the equipment, making it difficult to restart it.

A winterization concept should be implemented when developing the P&ID to deal with this issue. Contrary to the word *winterization*, it is not limited to winter time. Using a winterization tries to prevent the settling of remained fluid in a piece of equipment because of low

temperature, regardless of the season. Winterization will be discussed in more detail in Chapter 17.

If the equipment cannot be emptied, it should be checked to ensure that the remaining fluid does not create any problems for restarting the equipment. Restart after a shutdown is not always easy, and several steps may be taken to start the plant again.

If high-inertia equipment continues to run after emergency shutdown, this may cause a problem. For example, a large centrifugal compressor will be left with rotating impeller on for several minutes after an emergency shutdown. During P&ID development, an available system to continue the lubrication of the centrifugal compressor needs to be planned; otherwise, the compressor bearing may fail quickly.

To summarize, in developing the P&ID to cover the shutdown stage of a process plant, a good interlock system and winterization, among other project-specific items, should be implemented.

5.4.6 Inspection and Maintenance

Inspection, rejuvenation, and maintenance includes a wide range of activities that will be discussed more in detail in Chapter 8. Here, this topic is briefly covered. Because plant items are not eternal, they should be inspected, and if they need rejuvenation or maintenance, it should be performed.

Maintenance or rejuvenation can be triggered by any of three situations: after failure, based on inspection, and based on time intervals. The first one is termed *corrective maintenance*; the second one is *inspection-based preventive maintenance*; and the third is *time-based preventive maintenance*.

Therefore, two stages of a process plant are revealed: inspection and maintenance and the time that a component is out of operation because of maintenance. In this section, the inspection and maintenance are covered, and in the next section, the period of time an item is out of operation because of maintenance is discussed.

During P&ID development, provisions allowing easy inspection and maintenance should be included.

This stage of the process plant is applicable again to the equipment, control system, and utility system, and it needs to be determined which equipment or system needs maintenance at what frequency and how long it takes to maintain it. These questions can be answered quantitatively or qualitatively.

5.4.6.1 Quantitative Approach to Maintenance Requirement

Answering these questions needs a vast amount of knowledge, which is generally the territory of a mechanical integrity engineer. However, there are two important

Table 5.9 Some typical values of MTBF and MTTR for three items.

	MTBF (yr)	MTTR (hr)
Centrifugal pump	1–3	8–16
Centrifugal compressor	10–20	6–24
Shell and tube heat exchanger	10–20	16–72

MTBF, mean time between failure; MTTR, mean time to repair.

parameters: mean time between failure (MTBF) of the equipment and mean time to repair (MTTR) of the equipment. These two parameters specify the availability of a piece of equipment.

Table 5.9 outlines some typical values of MTBF and MTTR for a few items. This table shows that, for example, a typical centrifugal pump will fail every 1 to 3 years, and its maintenance may take between 8 and 16 hours.

It is important to know that there is no unique value for MTBF and MTTR of centrifugal pumps. The parameters depend on the pump manufacturer, the type of process plant, the environmental parameters (ambient temperature, air humidity, etc.), and also the skill level of personnel. Each company may have their own database to collect the data and estimate MTBF and MTTR of their equipment and instruments.

5.4.6.2 Qualitative Approach to Maintenance Requirement

During P&ID development, a qualitative approach to the maintenance requirement is needed as it provides a bigger perspective encompassing *all* equipment, especially for the expensive ones. Here is a question we try to answer qualitatively: what equipment needs more maintenance?

In process plants, inspection is done on any or all of the features of a piece of equipment, for example, process goal, equipment functionality, and equipment integrity. This means that during inspection, we check if the process is running smoothly, if a piece of equipment is functional, or if there is no breakage in the equipment. The answer can be provided for each goal of inspection: process, equipment functionality, and equipment integrity.

For the processes that are based on a more probabilistic phenomenon, they may need more frequent inspections. The examples are burners and vessels that contain media like sand filters.

A peep hole is generally provided to observe the existence of flames in burners and sight glasses on the vessels with media.

For equipment functionality and integrity, the answer depends on the equipment and the process condition. It means some equipment inherently needs more maintenance attention, and some process conditions

make the equipment needier for maintenance. These two components are discussed next.

The equipment that are static generally need less maintenance. Among nonstatic equipment (i.e. dynamic equipment), the ones with linear (reciprocating) movements may need more maintenance attention than the ones with rotary movements (Figure 5.25).

Where there is a rotating shaft in a piece of equipment, the high rotational speed shafts (high revolutions per minute [RPMs]) may need more maintenance attention than low RPM shafts. Pieces of equipment that have tight clearances may need more inspection and maintenance. This is especially true if they are being used in services that are not clean.

When it comes to process and process conditions, the equipment that works in very high or very low temperatures or pressures may need more maintenance attention. The equipment that processes *non-innocent fluids* (i.e. highly acidic, precipitating, scaling, fouling, or any other aggressive fluid) may need more maintenance attention.

5.4.7 Operability in Absence of One Item

The designer needs to decide the repercussions of equipment loss, which means in the absence of a piece of equipment, it needs to be decided what will happen to the rest of unit or plant. The wide range of answers and decisions include:

- 1) **Do nothing!** In this case, the piece of equipment, unit, or even plant should shut down in the absence of a piece of equipment or instrument. This option should be avoided. Sometimes it is inevitable when a piece of equipment of interest is the main or one of the main pieces of equipment of the plant.
- 2) **Accumulation of fluid in middle containers.** In this solution, placing two containers with enough residence times upstream and downstream of the absent component help to prevent the absence of the component get “visible” by the rest of plant. In this solution, the upstream container allows the accumulation of fluid, and the downstream container provides flow for the downstream units.
- 3) **Redirecting the in-flow to a “reservoir” for later usage.** In this solution, the feed to the equipment can be redirected to a temporary reservoir (like waste tank or pond) to be processed later by returning it back to the system. Usually this is solution is not available for gases or vapors.
- 4) **Redirecting the in-flow to an “ultimate disposal” system.** This solution is the same as previous one, but the flow sent to the external reservoir cannot be returned. The feed to the equipment can redirected

to a waste-receiving system, like a flare system. This option can be considered if the preceding option is not doable. The previous option is definitely a better option because valuable materials are not lost.

- 5) **Bypassing the absent item.** The feed to the equipment can be bypassed temporarily with marginal impact on the operation of the system, like bypassing a trim heater if being off-temperature does not hurt the plant for a short time. There are some cases that is decided to bypass the equipment or unit when it is out of operation. This can be done if the lack of equipment or unit does not affect the process in the short term.
- 6) **The nearly “similar” item in parallel.** A nearly similar system in parallel can take care of the flow that used to go to the absent system but not necessarily with the same quality. One example is having a manual throttling valve (e.g. globe valve) in a bypass loop of a control valve. The other example is placing a bypass pipe for a pressure safety valve (PSV) together with a pressure gauge (or pressure gauge point) and a globe valve. In the case of pulling the PSV out of operation, an operator will act as a PSV by monitoring the pressure of the container and being prepared to open the valve if it is needed.
- 7) **The exact “similar” item in parallel.** A parallel, exact replica as spare system can take care of the flow that used to go to the absent system. This is the most expensive option. The examples are all spare pumps or spare heat exchangers (in very fouling services). Spare equipment are very common for fluid-moving equipment as usually the pumps and compressors cannot be handled otherwise. One important example is having two fire pumps in parallel with two different types of drives (i.e. one electromotor and the other one a diesel drive pump). The spare can be in different forms.

In Table 5.10, the schematics of these options in the P&ID are shown.

5.4.8 Provision for the Future

The other concept that may affect the development of the P&IDs are provisions for the future. The future arrangement of a plant is not necessarily similar to the current arrangement of plant because the future of a market is not always foreseeable, or if it is foreseeable, it is not economically justifiable to incorporate it into the current plant design. However, to minimize the cost of rearrangement of a plant in the future, some items can be placed in the plant design and the P&ID. Therefore, some “footprints” of future on a P&ID may be seen; however, not all plants consider the future.

Table 5.10 Options for dealing with lack of a component.

Option	Schematic	P&ID example
1 The lack of item does not generate any upset in the rest of plant or whole plant should be shut down		
2 The upstream tank stores the flow in and the downstream tank provides flow out for a short period of time		
3 Redirecting the in-flow to a reservoir for later sending it back		
4 The flow-in is sent permanently for ultimate disposal and the stream will be wasted		
5 Bypassing the absent item		
6 The similar item in parallel		
7 The exact replica in parallel		

The future programs can be a planned or unplanned program. A planned change in a process plant could be scheduled or unscheduled and yet still be a *foreseeable* change.

Different words are used to describe a planned future program. *Revamping* is a general term that refers to any change to a plant. *Upgrading* means implementing new

inventions or practices to decrease the operating cost or increase the quality of product(s). *Retrofitting* means adding new features to the plant to produce additional products, better products, or the same product at lower cost or with less safety concerns. Retrofitting can be needed because of a change in feed type. *Expansion* means increasing a plant capacity to match the new

requirements of the market. Expansion can also be scheduled on a specific date in future or unscheduled whenever the market picks up and needs more of a particular product. Expansion can be done by adding a new “train” or adding or upsizing several (hopefully not *all*) pieces of equipment. In this context, a *train* is a similar plant with the same capacity next to the original plant and which may share some of the same resources as the original plant. A plant can be designed and built based on the training concept because of a limited available budget or other reasons. Different trains of a plant may have cross piping to each other to increase the reliability of the plant (Figure 5.26).

It is very common that multiple trains of a plant share the same utility system (including utility generation plant and utility distribution and collection network). In such cases, all the utility generation plant (e.g. boilers, coolers, potable treatment package, utility networks) should be upsized, and the utility network be ready and extended through the future train. The upsized pipes may have plenty of blind flanges at different locations of utility networks and process headers.

An unplanned future program is categorized in all debottleneck activities, rehabilitation, or optimization. No design is perfect, and to bring a plant design even near perfection may lengthen the project duration beyond economically accepted measures. Therefore, almost always, the designers do their best to design a plant within the allotted time for the design, and they leave the additional fine-tuning for later and after start-up of the plant. There is a better chance to fine-tune a design during the operations of a plant rather than before its start-up. However, a specific debottleneck activity is not always needed; there are cases in which the plant owners are comfortable with the design and they do not do any debottlenecking on the plant currently operating.

Rehabilitation is one common example of an unplanned future program. Rehabilitation means replacing old equipment with new equipment or refurbishing old equipment for the purpose of decreasing the operating cost or increasing the quality of products.

Unplanned future programs do not have any footprint on a P&ID until they are initiated.

5.5 Conflicting Check and Merging Opportunities Check

This step has two substeps: checking if a conflict is result of the functionality of added items and checking if some of added items can be merged into one item to save money.

5.5.1 Conflict Check

This step is making sure that no added element within one stage of the plant’s life cycle jeopardizes another item’s goal. For example, adding a bypass with a manual block valve for a safety-related switching valve for the purpose of making the plant operational when the switching valve is out for maintenance (point 5) could jeopardize the intent of the switching valve’s operation in a safety instrumented system (the bypass could left open and therefore is a safety flaw).

5.5.2 Merging Opportunities Check

Merging can be done in two different levels: merging different components including equipment, containers, pipes, or instrument or merging different items on one specific component. The former is the opportunity to use one pump, instead of two pumps, to handle two services. The latter is the opportunity to use one nozzle on a tank as an overflow nozzle *and* vent nozzle.

This step is deciding whether the added items or components can be merged with each other. This is checking if a single shared item or component can address multiple requirements of the different plant life cycles or not.

The quick answer to this question is to merge the items or components and use the time-shared item as much as possible to save money. In simpler terms, for the sake of saving capital and operational costs, the required items or components should be shared. It is better to put more than one duty as much as possible on the shoulders of each added item or components. However, shared items are more difficult to fit into the design, may be confusing to operators, are more prone to cross-contamination, and can generate a big shutdown caused by a small failure (redundancy issue).

A common example of merging multiple components in a plant happens in batches and intermittent operations. If you have a pump operating in an intermittent mode, you may be able to put another duty on the shoulder of this pump when it is freed from its main duty. It means that some components in such operations can be used for different duties in different time spans.

Merging items are doable when one item is functioning in a time phase other than that of the similar item.

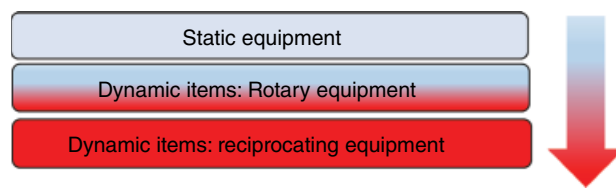


Figure 5.25 Necessity of inspection and maintenance for equipment.

For example, if an item was added only for the ease of maintenance, another duty may be placed on its shoulders during normal operation.

Later in this book, each piece of equipment and opportunities for merging them are discussed.

5.6 Dealing with Common Challenges in P&ID Development

During the development of a P&ID there are occasionally some challenges to find a better option among the available options. Sometimes these challenges are in the designer's mind and are resolved easily, but sometimes a challenge can be the subject of heated debates between stakeholders. Following are a few listed and discussed.

- **“Should I add this item or not?”**

The components and items should be added to give the operator enough flexibility. A plant with not enough resources is difficult to operate, and it is also the case for a plant with more than enough pipe circuits, control valves, alarms, and SIS actions. For example, a plant with too many alarms will overload

the operator, which results in operators losing a sense of urgency in the case of an alarm (Figure 5.27).

However, the designer should be careful of not falling in the trap of “adding does not hurt!” This is a popular statement when P&ID developers try to bypass the complete evaluation of the need for an item in the system and placing it in the system regardless. However, although adding an item might not increase the capital cost of the project (if it is small and inexpensive), it will increase the operating cost because of the required inspection, maintenance, probable utility or chemical usage, and so on. In addition to that, any new item in the system is an opportunity for mistakes, cross contamination, and leaks.

- **“Based on my past experience...”**

The inherent creativity required in developing P&IDs may become to hinder, if for every single case one refers to past experience. Every past experience should be reevaluated and tailored before being applied to new situations. Unlikely as it may seem, the “this is what has been done before” mentality is not the most efficient way to develop P&ID. On the other hand, technological innovations, availability of materials, quality of raw material, and the required quality of products,

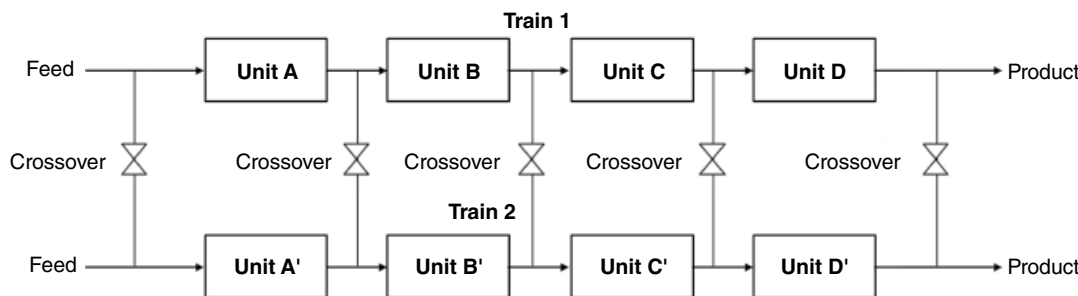


Figure 5.26 A “train” in a plant.

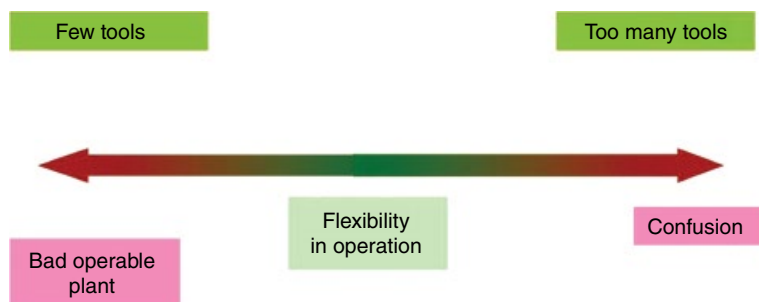


Figure 5.27 The “sweet spot” for providing items for a plant.

capacity of the system, and ambient temperatures and pressures will most likely differ for each project. In the world of P&ID development, a previously effective method may be entirely ineffective in a different project, and a method proven useless in the past may work perfectly for a new project.

- **“Based on chat I had with my friend last night in bar...”**

Conversations in informal meeting tend to be imprecise. Although it cannot be said that everything verbal is wrong, the facts conveyed in informal meetings and verbally should be checked against other facts to be reliable.

- **“Should I add it here on P&ID, or will it be captured in other documents?”**

The P&ID is supposed to be a common document for a few different groups. Incompleteness is an inherent feature of it. Furthermore, the P&ID is supposed to be kept in the plant for operators. If it is crowded, it cannot be used easily.

All process equipment should be shown on the P&IDs. Sometimes, nonprocess-related P&IDs (like gearboxes, lubrication systems, etc.) are also shown on the main P&ID or on auxiliary P&IDs. If they are not on the P&IDs, their details can be found in vendor documents.

All pipes and pipe appurtenances, except bend and elbows, are shown on the P&IDs. Flanges are depicted

if there is a specific reason for them. The piping items that are not on P&ID can be found on piping models.

Instrumentation and control systems are debatable. The three main elements of a handling system in a plant are: regulatory control system (BPCS), alarming system, and SIS. Almost everyone is clear about the items out of the BPCS that should be shown on P&IDs. They are mainly elements of control loops. For alarming system, the same clarity exists. The main debate is usually on SIS system: “Down to which level of detail the safety interlock loops should be shown on the P&IDs?” Companies do it differently.

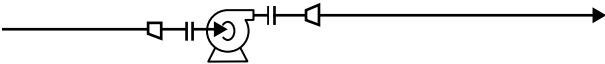
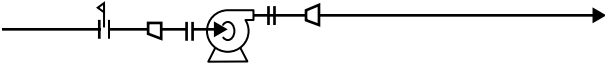
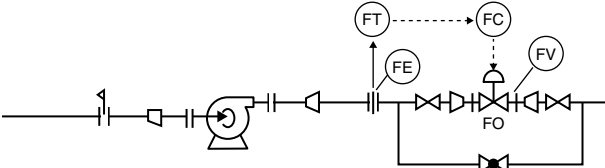
- **“Should I show it on the main body of the P&ID as a schematic, or I can capture it in the Note area?”**

A P&ID is a pictorial diagram, and most of what should be accessible for use should retain a schematic shape as much as possible.

5.7 Example: Development of P&ID of a Typical Pump

The following is an example of developing P&ID. If you are not familiar with the symbols here, do not worry. Revisit this example after studying Chapter 10. The points that have to be considered for developing a pump P&ID are shown in Table 5.11.

Table 5.11 Development of a centrifugal pump P&ID.

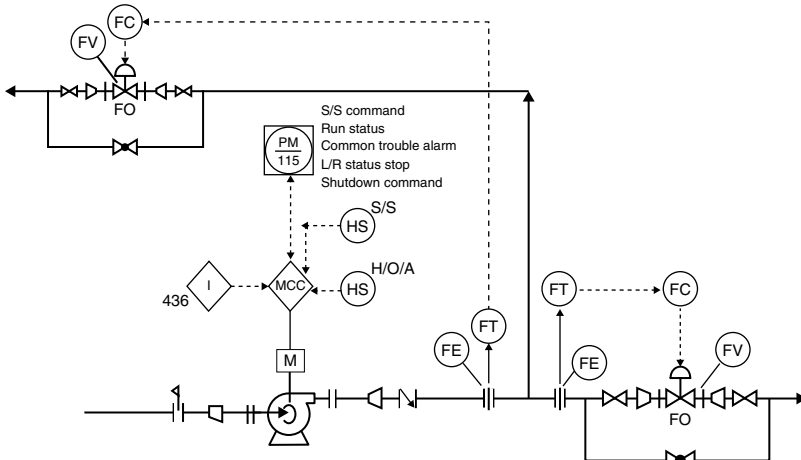
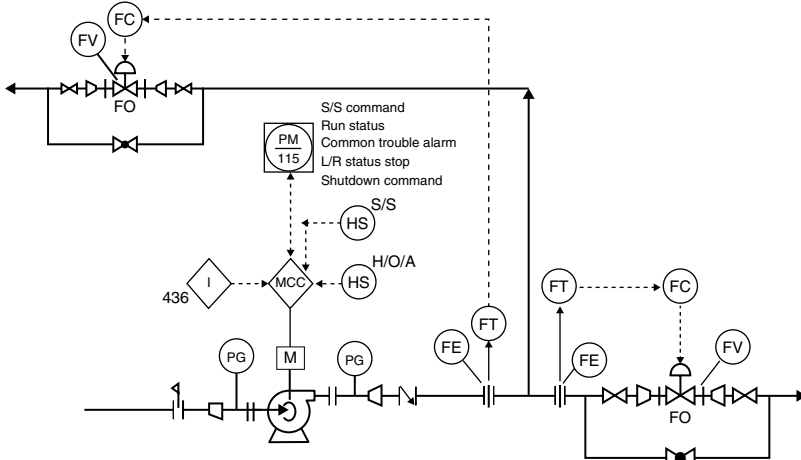
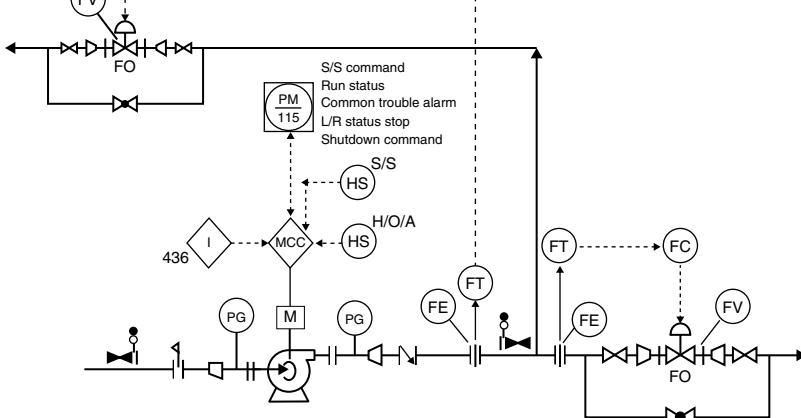
Case	P&ID
<i>Normal operation</i>	
Putting the pump callout with the required information on top of P&ID sheet	Callout should be put on P&ID
Placing reducer or expander to match suction and discharge side of the pump, if needed (mention using a top-flat eccentric reducer at connection)	
Adding permanent strainer to prevent damage to the pump	
Making sure the suction pressure (and temperature) is enough. This reflects the sensitivity of centrifugal pumps toward NPSH	It needs some calculations. The impact on P&ID could be seen on suitable LLLL in upstream container of pump.
Showing pump BPCS for capacity control of pump	

(Continued)

Table 5.11 (Continued)

Case	P&ID
Adding a pump's driver control	
Nonroutine condition	
Considering a temporary strainer (commissioning)	
Adding a nonreturning valve in case of reverse flow	
Using the minimum flow line on a discharge line with a control valve to protect the pump from flows lower than minimum flow of the pump	

Table 5.11 (Continued)

Case	P&ID
Showing the pump SIS or alarming system to protect the pump from an abnormal condition (such as seal leakage monitoring system)	
Maintenance and inspection Adding a pressure gauge in discharge or suction	
Adding a block valve in suction and discharge lines (such as a gate valve) to isolate the pump during the maintenance	

(Continued)

Table 5.11 (Continued)

Case	P&ID
Consider vent and drain valves in pump suction and discharge sides' pump casing	<p>The diagram shows a pump system with a central pump (M) driven by a motor (M). The pump is connected to a suction line and a discharge line. The suction line has a vent valve (FV) and a drain valve (FO). The discharge line has a vent valve (FV) and a drain valve (FO). The pump is controlled by a PLC (PM 115) which receives signals from various sensors and actuators. The PLC outputs commands for S/S (Safety Instrumented System), Run status, Common trouble alarm, L/R status stop, and Shutdown command. The PLC is also connected to a H/O/A (High/Low/Alarm) system. The diagram includes various instrumentation symbols such as pressure gauges (PG), flow transmitters (FT), and level transmitters (L/T). The pump is labeled with the number 436.</p>
Consider piping spool pieces for assembling or dismantling purposes	It is already created and exists
Pump insulation for personal protection	Service temperature is 40°C and no need for personal protection insulation
Production interruption	
Define the pump sparing philosophy	Based on RAM analysis, a second pump with the same arrangement is added to the current sketch (2 × 100%)

BPCS, basic process control system; NPSH, net positive suction head; SIS, Safety Instrumented System; RAM, Reliability, Availability and Maintainability.

Part II

Pipes and Equipment

In general, all process industries have three main elements: equipment, utility generation and networks, and instrumentation/control systems.

The second element of the process industry is the utility generation plant, along with the utility distribution and collection network.



The first element is the string of equipment.

In each plant, pieces of equipment are tied together to convert raw material(s) to product(s). In this context, the equipment could be pipes, vessels, tanks, pumps, heat exchangers, etc.

These pieces of equipment generally need external “help,” or utilities, to do their duty. For example, a pump needs electricity to operate, so in this case, electricity is considered as a utility. A heat exchanger may need a heat stream, like steam, to change the temperature of a process stream. In this example, “steam” is a utility.

All these utilities need to be generated in an auxiliary plant near the main process plant; this utility plant is called a utility generation plant. A utility generation plant is in nature the same as a process plant, but it only produces utilities.

There may be different types of utilities in each process plant, including electricity, steam, utility water, instrument air, utility air, cooling water, etc.

The generated utility in the utility generation plant should be distributed to different users or equipment. In some cases, the “used utility” needs to be collected to save some money by recycling the utility by converting it to a “fresh utility” for other reasons. In such cases, a collection network is also needed.

The third element of the process plant is the instrumentation and control system. This third element works like the nervous system in the human body in that it monitors different locations of the plant and controls their operation.

Therefore in Part 2 we are going to cover “pipes and equipment”. Parts 3 and 4 of this book are devoted to “Instrumentation and control” and “utilities”

Equipment

There are hundreds of different types of equipment in the different process industries. The types of equipment in an edible oil processing plant could be different from the equipment in an oil refinery; the equipment in a mineral processing plant could be different from the equipment in a waste water treatment plant. However, there are five main groups of equipment that are common in almost all process plants:

- 1) Fluid conductors: pipes, tubes, ducts
- 2) Valves
- 3) Fluid movers: pumps, compressors, etc.
- 4) Containers: tanks, vessels
- 5) Heat exchange equipment: heat exchangers, furnaces

In this book, we focus on these items because by learning about them, the majority of items will be covered. Additionally, it will give you enough fundamental knowledge to understand other equipment in your specific industry.

For each of the above “items” these aspects will be discussed.

Their function and the specific features of them, their identifiers on P&IDs, Their naming, their different types and the general rules to choose amongst them, the different arrangements of them, series or parallel, opportunities to merge similar items to save money, and the requirements of items for all the stages of its life in a process plant.

Part 2, has seven chapters from Chapters 6–12.

In Chapter 6 we will cover pipes and briefly bulk solid handling systems.

Chapter 7 covers valves, either manual or automatic. However the role of automatic valves in process plants will be discussed in Chapter 13.

Chapter 8 talks about the stages of an elements’ life: inspection and maintenance. In Chapter 5 four stages of elements were briefly discussed: normal operation, non-routine operation, inspection and maintenance, and finally the absence of an element. While other life stages of elements are covered when discussing each element the inspection and maintenance stage needs specific attention. Chapter 8 is a chapter devoted to this concept.

In Chapter 9 we will cover containers, including tanks and vessels.

Fluid movers will be discussed in Chapter 10. Fluid movers include pumps, fans, blowers, and compressors.

Chapter 11 talks about different heat transfer equipment. These include heat exchangers and furnaces.

Chapter 12 covers pressure release devices.

6

Pipes

6.1 Fluid Conductors: Pipes, Tubes, and Ducts

In process industries, to transfer fluid from point A to point B (e.g. equipment A to equipment B), a fluid conductor is needed. Pipe is the most common type of fluid conductor as it can transfer liquid, vapor, gas, or flowable solid. The fabrication process of pipes is different from tubes. For a small fluid flow, tubes are used, while for low-pressure gases or air, ducts. The other fluid conductors are channels, chutes, and so on (Table 6.1).

It is not difficult to transfer fluids, liquid, gases, and vapor, but flowable solids is. The word *flowable* is used for a noncontinuous version of solids. These solids could be in the form of granules, pellets, flake, powder, beads, chunk, and so on. They could be sugar or tomato in food industries or ore in mineral processing. Usually the smaller the solid particles, the easier they are to transport. To transfer a stream of iron ore in chunk form, we may need a bucket elevator rather than piping. Chutes are used for vertical or downward transfer of solids and semisolid materials. To summarize, transferring solids are generally done by equipment rather than by a simple conductor.

In this chapter, the main focus is on pipes.

6.2 Pipe Identifiers

As discussed in Chapter 4, the identifiers of pipes on P&IDs are pipe symbols, pipe tags, and stream names on the off-page connectors.

6.2.1 Pipe Symbol

The symbol for pipe is a line (Table 6.2). On P&IDs it is preferable to show pipes as vertical or horizontal lines rather than oblique. There are two features related to lines: their thickness and their arrowhead. On P&IDs, the thickness of lines gives more information about the

pipe. A thick line represents a primary pipe, and a thin line means a nonprimary pipe. The decision whether a pipe is primary depends on the purpose of the plant. Generally, primary pipes are used in the main feed and product of the plant, while nonprimary pipes are for other streams. It is important to know that the thickness of the line does not necessarily specify the diameter of the pipe. A thin line could be a 2-inch pipe, a 6-inch pipe, or even a 20-inch pipe. However, primary pipes do generally tend to be large bore pipes.

The other feature of line for pipe symbols is the arrowhead. The arrowhead shows the direction of flow in pipes during the routine operation. However, the question is, in which situations should the arrowhead be shown? The general rule is that the arrowhead should be shown in two cases:

- 1) Where there is a change in the direction of the line (Figure 6.1).
- 2) In the inlet of equipment (Figure 6.2).

It is important to consider we generally do not put an arrowhead on the inlet of valves or instruments (Figure 6.3). Some companies also do not put arrowheads on short loops around process or instrument items (Figure 6.4).

There are cases in which a pipe has different flow directions during different phases of routine operation. Because the arrowhead shows the direction of flow during routine operation, normal operation could be one stage of routine operation. For example, a sand filter that works in a semicontinuous mode may have different operation phases: filtration, backwashing, and retention. Then there are pipes that carry flows in a different direction in different phases of operation. Thus bidirectional lines may exist (Figure 6.5).





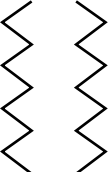
6.2.2 Pipe Tag

The second pipe identifier is a pipe tag. Pipe tags should be assigned to each piece of pipe and shown on P&IDs. The anatomy of pipe tag is specified by the project

Table 6.1 Different fluid conductors.

Conductor	Suitable services	Applicability
Pipe	Liquids, vapors, gases, or flowable solids	By default choice
Tube	Liquids, vapors, or gases	For smaller flow rate streams, for very high pressure streams
Hose	Liquids, vapors, or gases	For small to medium flow rate streams when portability is needed
Duct	Gases	Only for low-pressure gases
Channel	Innocent liquids	For innocent liquids (generally water)
Trench	Innocent liquids	For small flow rate streams when liquid is innocent
Chute	Flowable solids	For downward vertical transfer of bulk solids
Solid transfer equipment	Not very flowable solids	For not easily pushed solids

Table 6.2 Symbol for fluid conductors.

	P&ID symbol
Tube	Not generally shown
Hose	
Duct	
Channel	
Trench	
Chute	

guidelines. Different information packs in the pipe tag will be discussed in Section 6.4.

A sample pipe tag is shown in Figure 6.6.

All the tagged pipes will be eventually listed in a document called a line list or line designation table (LDT).

Three questions should be answered: Which flow conductors should be tagged? Which span of pipe route can be considered one piece of pipe? And, how is the pipe tag shown on a P&ID? These questions are answered here.

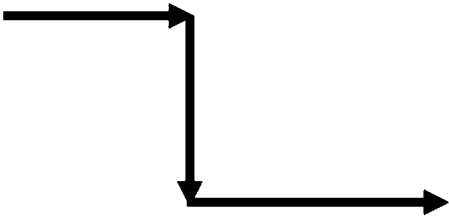


Figure 6.1 Showing an arrowhead on direction change.

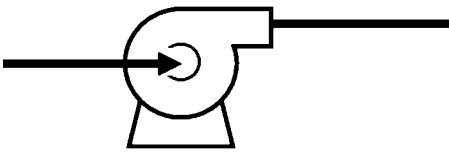


Figure 6.2 Showing an arrowhead on inlet of equipment.

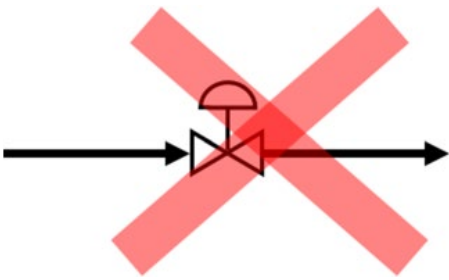


Figure 6.3 No arrowhead on inlet of valves and instruments.

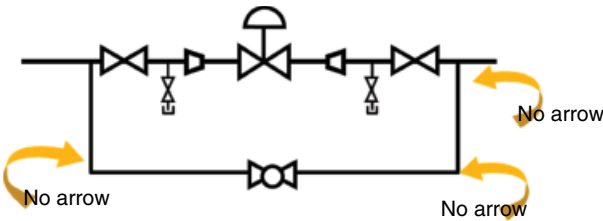


Figure 6.4 No arrowhead short bypasses.

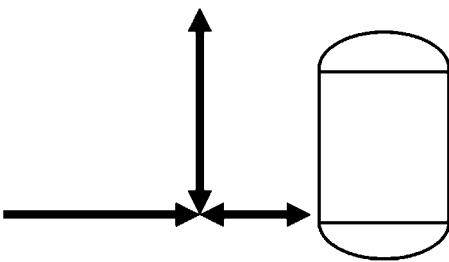
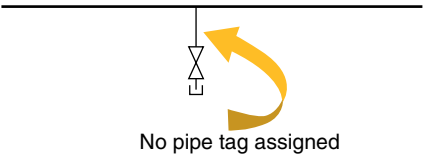
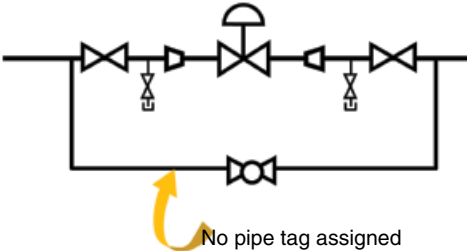


Figure 6.5 Bidirectional lines.

PW-04-210-AQ2-4" -ST (2-1/4")

Figure 6.6 A sample pipe tag.

Table 6.3 Two examples of pipes that may not assigned pipe tags.

Option	P&ID schematic
Short piece of pipes connecting process to drain or vent valve where the valves are plugged or blinded	
Bypass pipes around control valves	

6.2.2.1 Do All Pipes Need to be Tagged?

Not all flow conductors on a P&ID set are tagged. Generally speaking, only pipes are tagged. The question to answer is: Does it need to be hydrostatically (or pneumatically) tested before being put into service?

If the answer to this question is no, the flow conductor does not really need to be tagged unless it is long enough.

One main use of pipe tags is for documentation of hydrotesting, which checks the integrity of a piping circuit. After the performing the hydrotest, all the tag names of the pipes in a piping circuit that pass the test should be recorded. If a pipe does not carry a tag, it may be missed for hydrotesting, which is a big mistake.

Table 6.3 outlines some example of pipes that are not generally tagged. Hydrotesting is performed on a larger integrated pipe, so these untagged pipes will be still tested.

In another perspective, some stakeholders believe P&ID's in an operating plant shouldn't carry pipe tags at all! Plant operators claim that P&IDs are too crowded to be used during operation as some information should have been eliminated from the last revision of P&IDs before issuance for use, for example, the pipe tags because they are not useful during plant operation. However, pipe tags are a very critical piece of information in critical plants like nuclear power plants.

6.2.2.2 Which Span of Pipe Route can be Considered One Piece of Pipe?

It is a bad common practice to overly assign pipe tags. This results in too many additional tagging that further lengthens the line list and also leads to increase in the cost of pipe hydrotesting. However, while it is fine to excessively assign pipe tags, it should be mentioned

that this is not generally considered as an error in P&ID development.

A tag should be assigned to a pipe with clear start point and end point. The best start and end points are equipment. This means the equipment should be first identified before pipe tags are assigned. Sometimes it is difficult to find a border equipment. In such cases, another pipe could be considered the border for the pipe of interest, but this should be the middle of the pipe and not the end of pipe. Then it should be checked if the primary selected pipe span should be divided into pieces - depends on service spec - and a unique pipe tag assigned to each piece *or* all the selected pipes span can carry one pipe tag.

Therefore, it can be said that assigning a pipe tag has two steps.

Step 1: Pipe tags should be assigned to a pipe, regardless of length, between two pieces of equipment or the middle point of another pipe.

Step 2: Breaking the pipe down may be need to be done if temperature, pressure, or flowing fluid is changed.

It is important to know that change in temperature, pressure, or flowing fluid do not *always* warrant a change in pipe tag. This concept will be discussed in more detail in section 6.3.3.

6.2.2.3 How is the Pipe Tag Shown on a P&ID?

Pipe tags are generally shown as lines. Preferably, they are shown on the horizontal part of lines on a P&ID and above them.

As a general rule, repetition of a pipe tag is not allowed. Pipe tag is generally put once on each sheet of a P&ID and at the beginning of the pipe.

Figure 6.7 shows good and bad examples of indicating pipe tags.

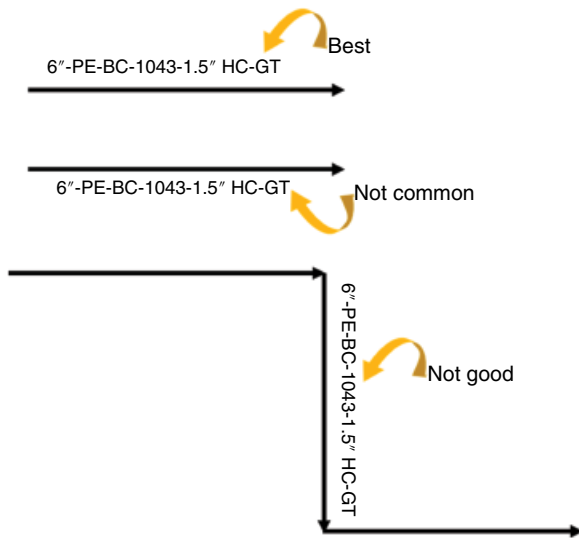


Figure 6.7 Good and bad examples of showing pipe tags.

6.2.3 Pipe Off-Page Connector

The third pipe identifier is pipe off-page connector. Wherever a pipe is introduced or removed from a P&ID sheet, it should carry an off-page connector with a set of information specified by the project.

Pipes on P&IDs are shown as lines. Because a plant is a large set of interconnected equipment and generally does not fit onto one sheet, there are some pipes that go from one P&ID sheet to another. Off-page connectors are arrow-shaped symbols that appear at the edge of P&ID sheets and show the continuity of each pipe. Off-page connectors are determined per project and are different within each company. Figures 6.8 and 6.9 show samples of an off-page connector for an outgoing pipe and an incoming pipe, respectively.

6.3 Pipe Tag Anatomy

A tag number bears many information about a pipe. The anatomy of a pipe tag varies from company to company; however, the following information can usually be found in a pipe tag:

6.3.1 Area or Project Number

Possibly the first component of a pipe tag is a number or letter that shows the area where a pipe is located. If a pipe goes from one area to another area, generally the area that is mentioned in the pipe tag is its origin. The area designation could be like 06, 21, or AB. The area number could be between one to three figures or letters.

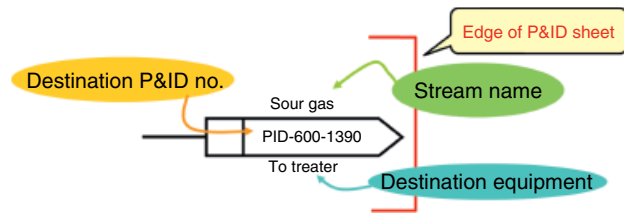


Figure 6.8 A leaving off-page connector for pipe.



Figure 6.9 An incoming off-page connector for pipe.

6.3.2 Commodity Acronym

The commodity name is basically the fluid that pipe is conveying. The acronym could be between two and three or more letters; however, using the acronym is not an arbitrary choice by the designer. For example, if the conveying fluid is water, the designer cannot arbitrarily choose to use the acronym of WAT, without consulting a commodity table.

Commodity acronyms are specific in each plant and project, hence the variation. For example, fire water can be FW or, in some companies, FWA. Table 6.4 lists those that are universally used among plants.

There are also some common rules in choosing an acronym for commodities. For example, H, M, and L at the beginning of utility codes mean high, medium and low. It is also common to see an S or R as the last letter, meaning supply and return, respectively. The two types of acronyms, S-ending acronyms and R-ending acronyms are widely used for utility pipes, which will be discussed in more detail.

6.3.3 Pipe Material Specification Code

In essence, this code is an acronym that specifies the construction material and thickness of the pipe of interest. It is simply called a pipe specification (spec) or pipe class. Each pipe spec is typically a string of two to four letters or numbers or a combination of them. Each company has its own pipe tag anatomy per pipe class. For example, a pipe class could be A2S. Companies try to develop pipe class anatomies that are more meaningful for the designer or reader of the P&ID.

Table 6.4 Commodity acronyms and their meaning.

Acronym	Fluid name
NG	Natural gas
FW	Fire water
IA	Instrument air
UA	Utility air
PA	Plant air
D	Drain
V	Vent
UW	Utility water
LPS	Low-pressure steam
MPS	Medium-pressure steam
HPS	High-pressure steam
HGS	Heating glycol supply
HGR	Heating glycol return
CWS	Cooling water supply
CWR	Cooling water return

Because the pipe spec is just an acronym for the pipe feature, one may expect another document that outlines the detail of each pipe class. This is the Piping Material Specification Table. There could be less than 20 to more than few hundreds pipe specs in a Piping Material Specification Table of a project or plant. Figure 6.10 shows one page of a piping specification table, which belongs to imaginary pipe spec of A0.

A Piping Material Specification Table is a large document, and thus it has a table of contents. The content are called a Piping Material Spec Summary, which is shown in Figure 6.11.

The duty of the designer is choosing the suitable pipe spec for each pipe in the P&ID. To do that, three pieces of information are needed: the name of flowing fluid, its required design temperature, and its required design pressure (Figure 6.12).

The designer may start with checking the piping spec summary to find the available specs for the commodity of interest. He/she may find two to three different suitable pipe specs suitable based on the commodity and the temperature range. Then the designer needs to pick one

Pipe Class:		A0				Date:	Apr. 28,17	
						Revision:	2	
						Sheet:	1	of 4
						By:	Moe T.	
Service:		Water, Air						
Base Material:		Carbon Steel/Low Temp Carbon Steel				Design Code:	ASME B31.3	
Internal Corrosion Allowance:		1.60 mm				Flange Rating:	ASME CLASS 150	
						Branch Table:	Note 1	
Material Group:		1.1				Inspection Class:	Note 2	
						Inspection Class:		
COMPONENT	NPS SIZE RANGE		SCH. OR RATING	STANDARD/CODE	MATERIAL DESCRIPTION AND SPECIFICATION	NOTES		
	FROM	TO						
Pipe, Seamless	½	1 ½		ASME B36.10	PIPE, SEAMLESS, ASTM A106 Gr. B, PE			
	2	3		ASME B36.10	PIPE, SEAMLESS, ASTM A106 Gr. B, BE TO B16.25			
	4	24		ASME B36.10	PIPE, SEAMLESS, ASTM A333 Gr. 3, BE TO B16.25			
Nipple	½	1 ½		ASME B36.10	PIPE, SEAMLESS, ASTM A106 Gr. B, SCH XXS, 4" LONG, PBE			
Swaged Nipple	½	1 ½		MSS SP-95	SWAGE NIPPLE, SEAMLESS, ASTM A234 Gr. WPB, PBE. WT TO MATCH PIPE			
	½	3		MSS SP-95	SWAGE NIPPLE, SEAMLESS, ASTM A234 Gr. WPB, BLE xPSE. WT TO MATCH PIPE			
Flanges	½	1 ½	CLASS 1500	ASME B16.5	SOCKET WELD FLANGE, RTJ, ASTM A105N			
	2	2 ½	CLASS 1500	ASME B16.5	WELDNECK FLANGE, RTJ, ASTM A105N			
	3	3	CLASS 900	ASME B16.5	WELDNECK FLANGE, RTJ, ASTM A105N			
	4	24	CLASS 900	ASME B16.5	WELDNECK FLANGE, RTJ, ASTM A350 Gr LF2 Cl. 1			
Blind Flange	½	2 ½	CLASS 1500	ASME B16.5	BLIND FLANGE, RTJ, ASTM A105N			
	3	3	CLASS 900	ASME B16.5	BLIND FLANGE, RTJ, ASTM A105N			
	4	24	CLASS 900	ASME B16.5	BLIND FLANGE, RTJ, ASTM A350 Gr. LF2 Cl. 1			
Orifice Flange	2	2 ½	CLASS 1500	ASME B16.36	WELDNECK FLANGE, RTJ, ASTM A105N, BE TO B16.25			
	3	3	CLASS 900	ASME B16.36	WELDNECK FLANGE, RTJ, ASTM A105N, BE TO B16.25			
	2	2 ½	CLASS 1500	ASME B16.36	WELDNECK FLANGE, RTJ, ASTM A350 Gr. LF2 Cl. 1, BE TO B16.25			
TTIP (90, 45 DEG ELBOW, EQUAL								

Figure 6.10 An excerpt of A0 pipe spec table.

Piping Spec	RATING	TEMP. RANGE (°C)	SERVICE
A0	150	–29 to 260	Lime slurry
A1	150	–29 to 93	Instrument/Utility Air
A2	150	–29 to 120	Lime, Mg(OH) ₂
A3	150	–46 to 0	Sweet H.C. Low Temp.
A4	150	–46 to 0	Sour H.C. Low Temp.
A5	150	0 to 60	Demineralized Water
A6	150	–46 TO 93	Potble water
A7	150	10 to 260	L.P.Steam/Cond/BD
A8	150	–41 to 149	Lube, Seal Oil
A9	150	–46 to 399	Sweet Hydrocarbon
A10	150	0 to 38	Steam/Condensate
B0	150	0 to 30	Chemical Injection
B1	150	–29 to 93	Hydrochloric acid
B2	150	0 to 93	Nitric Acid acid
B3	150	–29 to 93	NaOH < 40%
B4	150	–29 to 93	Sweet Gas, Wet CO ₂
B5	150	10 to 204	Sour Gas, Wet CO ₂ rich
B6	150	–29 to 260	Buried Fuel Gas
B7	150	–29 to 204	Produced Gas, Wet CO ₂
C0	150	–29 to 204	Glycol Tracing
C1	150	0 to 100	Caustic 50%
C2	150	0 to 93	Produced Salt Water
C3	150	0 to 93	Brackish Water
C4	300	–29 to 260	Heavy oil, Sweet
C5	300	–46 to 260	Heavy oil, Sour
C6	300	–46 to 204	Sour H.C. Low Temp.
C7	300	10 to 370	Steam/Cond/BFW/Glycol
C8	300	10 to 204	Sour Gas, Wet CO ₂ rich
C9	300	0 to 260	Produced Emulsion P/L
D0	300	–29 to 260	Sour Hydrocarbon
D1	300	–29 to 204	Sour Off Gas P/L
D2	300	0 to 100	Raw Water

Figure 6.11 A sample piping material spec summary.

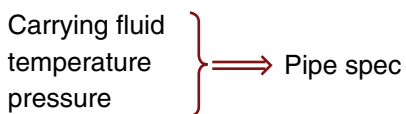


Figure 6.12 Functionality of pipe spec.

of them based on the pressure of service (Figure 6.12). Thus, the designer needs to go through the picked piping specs, study the design pressures and design temperatures, decide which one matches the required design pressure and design temperature of the pipe, and select that one as the suitable pipe spec.

The detailed procedure is as follows.

Consider the required design temperature of fluid in pipe: T0. Do interpolation between the commodity temperature–pressure range mentioned in the pipe spec of interest, (T1,P1) and (T2,P2), and find the corresponding pressure, P0. If P0 is between P1 and P2, pick the current pipe spec; otherwise go to a pipe spec with the same commodity but with higher pressure rating (Figure 6.13).

After selecting the piping spec, there is one important point that should be considered, but it is commonly overlooked. After selecting each piping spec, all the pages related to the selected piping spec should be read to check the limitation of the selected piping spec.

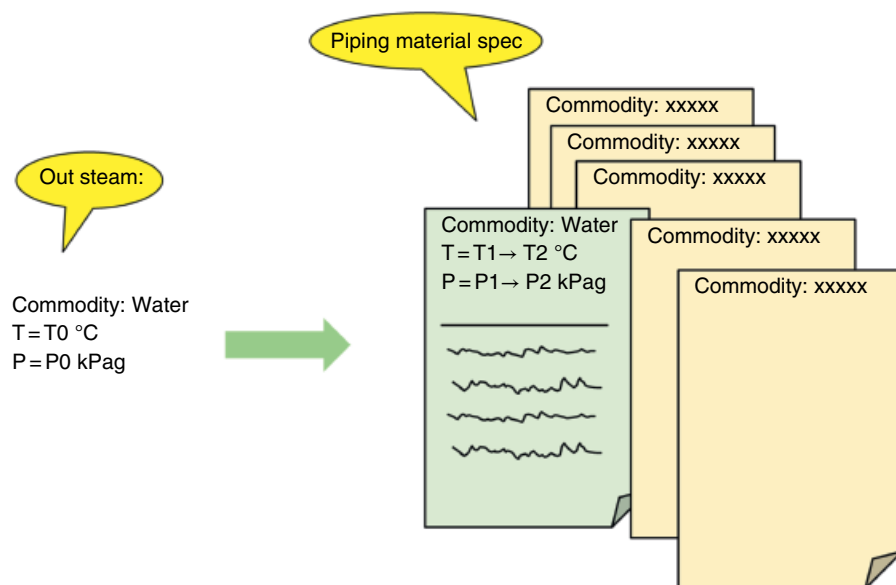


Figure 6.13 Specifying pipe spec.

The limitations could be the available pipe sizes or unavailability of some valves in the spec.

There are some cases that we are looking for something which is not available in the selected pipe spec.

There are at least four ways to deal with this issue:

- 1) Change the whole pipe spec by changing the interpretation of the piping spec commodity in a new, less radical way and move the fluid name to another less restrictive pipe spec.
- 2) Change the pipe spec to a less restrictive but still acceptable spec (a more rigid pipe spec).
- 3) Change the process design to obey the restrictions of the current pipe spec.
- 4) Ask the material group to update the current piping spec to cover the item of request.

It means there could be a specific piping spec for the utility water that has pipes only from 2 to 10 inches when you are looking for a 16-inch pipe in the plant.

The solutions are:

- 1) Is there another less restrictive pipe spec, for example, for raw water? Does it have 16-inch pipe in it? If the answer is yes, use it and change all the pipe specs to that one.
- 2) Is there another less restrictive, but more expensive pipe spec, for example, for potable water? Does it have 16-inch pipe in it? If the answer is yes, use it and put spec breaks for the 16-inch pipe to use this less restrictive but more expensive pipe spec.
- 3) Can you replace the 16-inch pipe with two 12-inch pipes in parallel to get over the limitation in the existing pipe spec?

- 4) As a last resort, ask the material group to update the piping spec table and extend their piping spec to cover the item of request (i.e. 16-inch pipe). This solution is not the best solution because companies are usually not willing to change their piping spec and changing the piping spec may take a few weeks to a few months.

The other limitation could be unavailability of some valves. In some piping specs, some specific valve types are not available. For example, in one piping spec, there could be no ball valve. So, if this is the specific pipe spec, it needs to be ensured that no ball valve is installed on the pipe. If there is a need to put a ball valve, the preceding solution can be used. An additional solution for valves is changing the valve to a similar valve with an actuator. Using this trick, the valves become beyond the piping specs and are transferred to the Instrumentation and Control group who may accept and approve the requested valve.

6.3.4 Pipe Size

The pipe size, or pipe diameter, to carry a fluid from point A to point B is already specified during the design phase of project. However, it is a good idea to have some practical understanding about these parameters.

Pipe size is generally mentioned as part of a pipe tag. However, the way that the pipe size is mentioned is different. Without going through different pipe standards, generally, in North America the pipe size is reported as nominal pipe size (NPS), which is an approximate size and not necessarily beyond the pipe diameter. NPS is generally stated in inches and an 18-inch pipe can be written as 18 in. or 18" pipe. Another way of stating

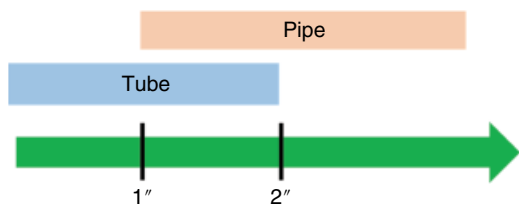


Figure 6.14 Availability of pipes and tubes.

pipe diameter is nominal diameter (DN) stated in millimeters (mm).

The NPS is at least less than 1 inch and goes up to more than 36 inches.

Less than 1 inch, the only available circular cross section fluid conductors are tubes, and more than 2 inches, pipes. The designer should specify the size because pipes and tubes are available in sizes between 1 and 2 inches (Figure 6.14).

The difference between pipes and tubes on a P&ID is the lack of pipe tag for tubes. A pipe tag is something only for pipes; tubes are not tagged.

There are some NPSs that are not common and are not easily found in the market. Some of the sizes are: 1 $\frac{1}{4}$ ", 2 $\frac{1}{2}$ ", 3 $\frac{1}{2}$ ", 4 $\frac{1}{2}$ ", 5", 7", 9", 14", 22", 26", 28", 32", and 34". Using these sizes is acceptable only if they are supposed to be used in long lengths. For example, a pipeline of a 14" pipe could be acceptable as a special order for a pipe manufacturing company that fabricates such lengths.

Pipelines tend to be on the higher side on NPSs, say, larger than 28". It totally makes sense because the capital cost for developing a pipeline is expensive, and to help recover the invested money faster, it is better to use large bore pipes as pipeline. However, there are some cases when using a small bore size pipe as pipeline is inevitable. For example, produced water is to be transferred from an oil-extraction site to a far disposal well. The pipeline size could be 6" or even 4" and depends on the flow rate. Making the pipe bore larger does not help the project financially.

Pipe sizes between 6" and 26" are common as main (or primary) piping in process plants. The nonprimary piping in process plants could be 6" to 2" pipes.

The 2-inch pipe is generally assumed to be the minimum practical pipe size within process plants for different reasons. For example, a 2" pipe is assumed as the minimum pipe size for gravity flow pipes. This is to make sure that the empty portion of the gravity flow pipe is not pressurized. If a pipe is supposed to be routed through a pipe rack, again its minimum pipe size is 2". A pipe smaller than 2" likely sags when it is laid down on a pipe rack because the pipe rack legs are placed in standard distances and the "weak" pipes may sag on pipe racks (Figure 6.15).

Oh, here's a 1" pipe that needs to go through a paperback; what we can do?

One solution is to put an "enlarger," increase the pipe size to 2", send it over the pipe rack, and, where it comes off of pipe rack, put a reducer and decrease the pipes size from 2" to the same 1" pipe. The increase of pipe size decreases the fluid velocity and it needs to be acceptable.

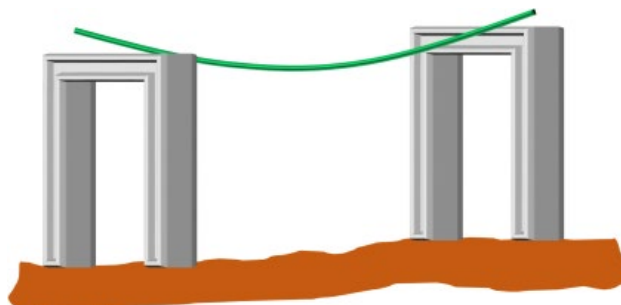


Figure 6.15 A smaller-than-2" pipe on pipe rack.

These tubes (bore size less than 1") are used as instrument air and steam tracing of pipes. Tube sizes of 1/8", 1/4", 3/8", and 1/2" are common in the instrumentation world. The tube sizes of 3/8", 1/2", and 3/4" are common as steam tracing tubes and also for pump seal arrangements.

6.3.5 Pipe Sequential Number

This is a sequence number that specifies every single pipe in a plant.

6.3.6 Other Pipe Tag Information

The following can also be included in the pipe tags. Some are discussed in other chapters of this book. Figures 6.16 and 6.17 show two examples of pipe tags.

- Insulation type and thickness (Chapter 17).
- Heat tracing type (Chapter 17).
- Whether the pipe is underground or aboveground.

PW-04-210-AQ2-4"-ST (2-1/4")

Commodity	Unit no.	Line no.	Pipe spec	NPS	Steam tracing	No. of tracers	Size of tracers
-----------	----------	----------	-----------	-----	---------------	----------------	-----------------

Figure 6.16 An example of the interpretation of pipe tags.

CD-AA-8"-2012-1"HC

Fluid name
Pipe spec
NPS
Line no.
Insulation thickness
Insulation type

Figure 6.17 Another example of the interpretation of pipe tag.

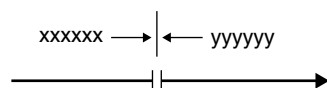


Figure 6.18 A border in a P&ID.

6.4 Pipes Crossing “Borders”

When or if a pipe crosses a border, this should be shown on the P&IDs (Figure 6.18). There are six different borders that a pipe can cross, and these should be depicted on the P&IDs.

- 1) Battery limit border: The battery limit refers to the border of the plant. It is a border around a plant that shows what belongs to the plant; those outside the border may belong to someone else. The battery limit border can be shown in Figure 6.19.
In the schematic, OSBL means outside battery limits and ISBL means inside battery limits.

- 2) Area border: Inside each battery limit, there may be a few different areas. Plants are generally divided into areas to make it easier to manage. There may also be technical reasons; by dividing a plant into areas, each area can be operated by a group of operators who receive training for that specific area. This is a more efficient way to operate a plant. In a mining plant, we may have areas 200, 300, 400, and 500. Whenever a pipe goes from, say, area 300 to 500, it should be depicted on the P&IDs. Typical symbology for such a border crossing is shown in Figure 6.20.

Even in some cases, pipe racks or pipe trenches are considered areas, and borders are depicted as seen in Figure 6.21.

There are some cases where a plant has different areas, but because these areas are operated by one group, the owner of the plant asks not to show the border between areas.

- 3) Building border: Equipment in a plant can be located inside or outside buildings. Whether a piece of equipment is located indoors or not is not generally mentioned, unless when/if a pipe goes from indoors to outdoors or vice versa. By looking at the building borders for a pipe, it can be determined if a given piece of equipment is indoors or outdoors (Figure 6.22).

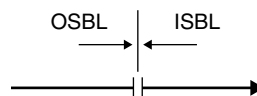


Figure 6.19 A battery border.

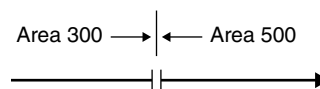


Figure 6.20 An area border.

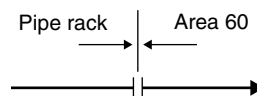


Figure 6.21 A specific example of an area border.

One way of showing building borders is as seen in Figure 6.23.

In this schematic, O/S means outside the building, and I/S means inside the building.

Why do we bother to show whether a piece of equipment is indoors or outdoors? It is because each of these two areas, indoors and outdoors, has specific features. For example, equipment in an outdoor area should be protected from harsh environments and ambient temperature changes. Indoor equipment is in a better environment, but it has limited accessibility because large cranes cannot get into the buildings and also a local building code may possibly be applicable for indoor equipment.

Generally speaking, everything is put in a plant outdoors because it is the less expensive option; however, there are some cases where some equipment should be mainly indoors for two reasons: if the equipment is sensitive to ambient temperature changes and/or if the equipment is operator intensive, meaning that it needs more attention from operators.

- 4) Work division border: This border is an important one during the construction phase of a plant wherein the majority of the construction is done by the construction group, managed by the owner or the engineering company. But there are some areas whose construction is done by the equipment or package vendor or manufacturer or licensor. In Figure 6.24, a work division border shows the area for which the construction should be done by the equipment vendor and not the plant owner or the engineering company.

It is common for work division borders to show a tie-point number whose symbol is generally a hexagon. The tie-point number facilitates the work of assemblers by telling them, “the pipe flange with tie-point 3 should be bolted to the flange of package 3”



Figure 6.22 Interpretation of a building border.

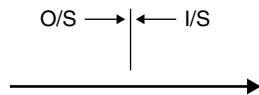


Figure 6.23 A building border.

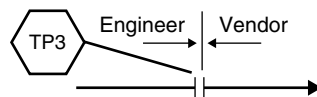


Figure 6.24 Work division border between two companies.

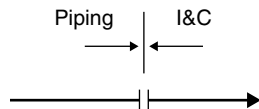


Figure 6.25 Work division border between two disciplines within a company.

Sometimes the division of work is also shown between two groups under one management team, as seen in Figure 6.25. For example, there could be a work division border on a pipe to show that the construction of a portion of the pipe should be done by the Piping group and the other portion by the Instrumentation and Control group.

This border can be seen frequently in instrument air pipe networks, which should be installed by the Piping group, except the tubing portion, which is under the Instrumentation and Control group during construction.

- 5) Ground border: When a pipe goes from underground to aboveground or vice versa, it should be depicted on the P&ID. Figure 6.26 shows ground border symbolology on a P&ID where A/G means aboveground and U/G means underground piping.

The requirements of underground piping are different than for aboveground piping. For example, underground piping may have a more robust coating system to protect it against external corrosion.

- 6) Pipe feature change border (other than pipe spec): Any change in any component of a pipe tag needs to be specified by the borders. This type of border is not a physical border; it is only for the purpose of highlighting the point on the pipe that a change starts. The

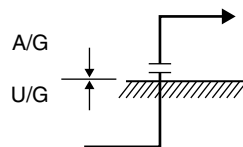


Figure 6.26 A ground border.

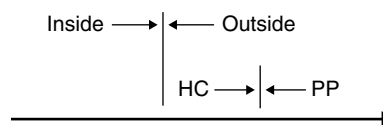


Figure 6.27 Insulation border together with building border.

change in a pipe spec is deeper and will be discussed in the next section.

For example, a pipe runs from outdoor to indoor and the insulation type changes from heat conservation (HC) type to personal protection (PP) type as shown in Figure 6.27. Different types of insulation will be discussed in Chapter 17.

The other types of these borders are the borders that show the extent of a specific pipe route (for example, a border for slope).

- 7) Spec break border: A spec break border is not a physical border, too. This “border” basically shows when the pipe spec changes. The pipe spec should be changed if the material or the type of pipe is changed. However, there are some cases that the pipe spec is changed for some nontechnical reasons, for example, when a specific required valve is not available in the existing pipe spec.

Figure 6.28 shows that the pipe spec changes from code BBA spec to code CBA spec. (For the type of material, refer to the piping spec booklet.) And by changing the pipe spec, the pipe tag is changed and pipe tags within the spec break are different. In Figure 6.28 two different pipe tags on each side of the spec break is seen.

6.4.1 Implementing Spec Break

As mentioned previously, the piping spec may need to be changed if the commodity (flowing fluid), temperature, or pressure is changed.

Figure 6.28 Pipe spec border and its effect on the pipe tag.

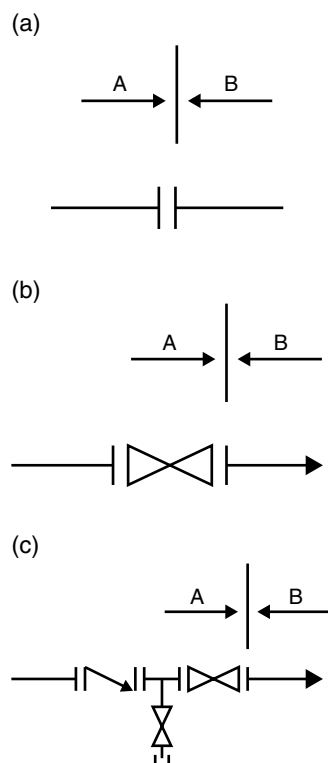
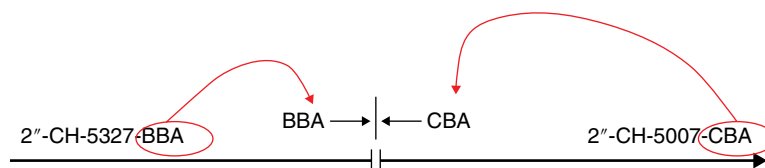


Figure 6.29 (a–c) Pipe spec break border systems.

But is this spec break only a nonphysical border or does it have some representation on a pipe like a flange? In a majority of border cases, only a flange is enough on a border, but in a spec break, the border could be more complicated. There are at least three different types of borders for spec break: flange (Figure 6.29a), blocking valve (Figure 6.29b), and blocking valve-check valve (Figure 6.29c).

A process engineer decides about the type of spec border based on judgment or consultation with the project documents. As a rule of thumb, a process engineer decides by default to use a flange for spec break. But if the spec is changed to a robust spec, the designer may choose to use a blocking valve or even blocking valve-check valve combination for spec break border.

For Figures 6.29b and c, the question that arises is which pipe spec should cover the border system (valves)? Or where should the pipe spec border be, on the right side or left side of the valves (Figure 6.30)?

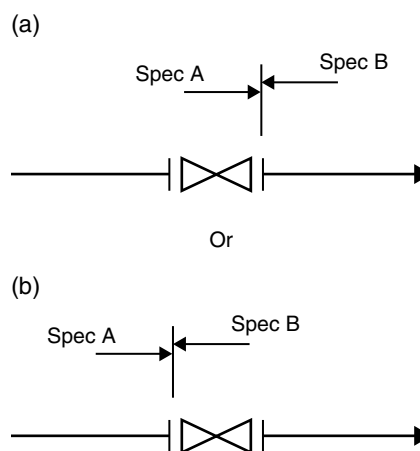


Figure 6.30 (a, b) Options for placing a pipe spec border on the border system.

The common practice is to always cover the valves with the more robust pipe spec. For example, in Figure 6.30, the designer should investigate if spec A or spec B is more robust. The more robust one should be included in the P&ID.

But the question is how we can recognize which is more robust or less robust. This is not always easy, so it is recommended to consult with a piping material engineer on the project.

Figure 6.31 shows a pipe spec border at the middle of a pipe, which is wrong because it should be at least on a

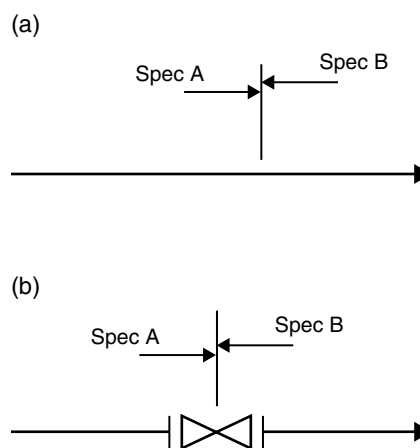


Figure 6.31 (a, b) Mistakes in placing a pipe spec border.

flange. It also shows a pipe spec border at the middle of a valve. Do we want to have these manufactured?

6.4.2 Reasons for a Spec Break

As it was mentioned, the pipe spec may need to be changed if the type of flowing fluid, the temperature, or the pressure of flowing fluid is changed. These changes together with other potential reasons for a change are listed in Table 6.5.

The question may arise: will the natural gradual pressure decreases in the pipe along the flow may cause the need for a spec break from high pressure to low pressure somewhere along the pipes? The answer to this is not generally so. During the normal operation of a plant, if the flows are not so high and the pressure drops are mild, then the designer does not need to change the pipe spec. However, this general rule has at least one exception; in blowdown pipes, the pipe spec may need to be changed for two reasons: the very high flow rate in blowdown pipes that creates a high pressure drop and the long length of blowdown pipes. These two makes the change in pipe spec from high pressure to low pressure economically justifiable.

The pipe spec break border is placed by default at the first point of temperature, pressure, or commodity change that warrants the pipe spec change.

For example, in Figure 6.32, a control valve severely dropped the pressure that the downstream piping could be with a less stringent pipe spec. The upstream pipe

Table 6.5 Three reasons for spec break.

Option	Decrease	Increase
Fluid change	Addition of new fluid to the flow of pipe	
Pressure	Partially closed valve or control valve or restriction orifice (RO) on the pipe	Pump or compressor on the pipe
Temperature	Heat exchanger(cooler) on the pipe	Heat exchanger (heater) or fired heater on the pipe
	Addition of cold fluid to the pipe flow	Addition of hot fluid to the pipe flow

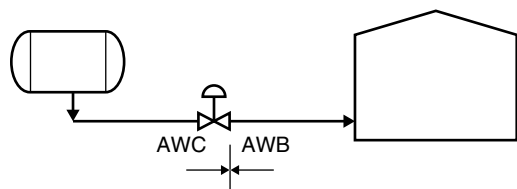


Figure 6.32 Pipe spec break because of a severe pressure drop by a control valve.

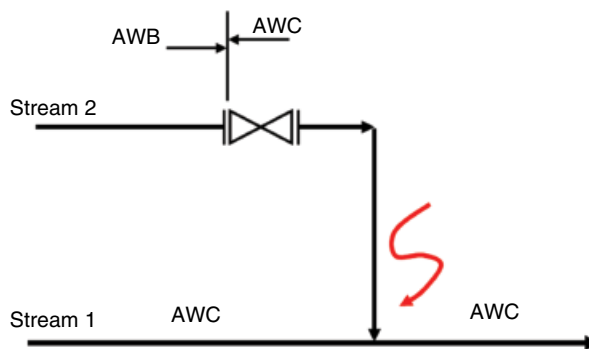


Figure 6.33 Pipe spec break because of a change in commodity.

spec is AWC, and the less stringent, AWB. The spec break border is placed right on the pressure change point and also because AWC is the more exacting spec, the valve is covered with AWC.

However, there are some cases that the concept of an extension of spec break to upstream should be applied. What this means is that under certain conditions the pipe spec break border is shifted to somewhere (but not very far from) upstream of the change point.

In another example, in Figure 6.33, stream 1 is a water stream and stream 2 is an acid that is injected to the stream 1. AWB is the acid resistant pipe spec, and AWC is the water pipe spec. Here instead of placing a spec break border on the arrow-pointed point, which is the point that the commodity changed, we placed the spec break border on upstream and on the first blocking valve of the acid pipe.

Table 6.6 is outlines different cases of a pipe spec break extension.

6.5 Goal of Piping

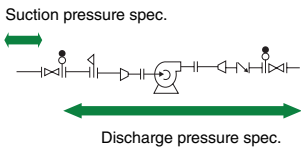
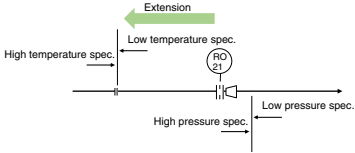
The goal of piping is transferring a fluid from one point to the other point in the required magnitude or flow rate (volumetric flow rate, mass flow rate).

However this goal will not be achieved until these requirements are satisfied:

- Pipe connection between points A and B.
- Flow goes in the right direction from point A to point B with the predetermined flow rate.
- The fluid is delivered to the downstream system (destination, or point B) at the required pressure.

Generally these requirements are checked during the design of the upstream and downstream equipment of the pipe. There are, however, cases that can be missed because they are considered so simple that calculations are not needed.

Table 6.6 Concept of a pipe spec break extension.

Changed parameter	Can the parameter penetrate to upstream?	Extension of spec break to upstream	Example
Fluid type	The changed fluid does not transfer to upstream point unless there are pressure fluctuations.	When a nonaggressive fluid changed to an aggressive fluid	The injected acid to a water stream may momentarily go upstream because of the pressure fluctuation Severe material pipe spec is extended up to the first blocking valve on the water pipe
Pressure	The pressure does not transfer to a high-pressure upstream point	Rare; there are cases, however, that pressure can go upstream	The idle pump can be pressurized by the operating pump in operating spare pumps in parallel  <p>High-pressure pipe spec is in pump discharge pipes but are extended up to the first blocking valve on the suction side pipe</p>
Temperature	The changed temperature can go upstream through conduction or convection	For metallic pipes when the temperature drop is huge	Temperature drop after RO on a depressurization pipe  <p>Low temperature pipe spec starts from 600 to 900 mm upstream of the RO.</p>

RO, restriction orifice.

Sample tube with a high-viscosity oil

In a project, a few sampling points are provided to take a sample from the oil accumulated in a water tank. Later, a client noticed there was no room near the tank for a sampling station. The sampling station and all the sampling tubes were placed far from the tank, which caused an additional 200m to the distance of the sampling tubes. The P&ID was updated based on this change, but later during operation, operators realized that high-viscosity oil cannot get through the long narrow tube.

After transferring fluid from point A to point B, point A should be able to manage the lack of fluid, and point B should be able to manage the accumulation of fluid and pressurization. The various ways to deal with this issue will be discussed in Chapter 9.

The following sections discuss the requirements of the piping.

6.5.1 Magnitude of Flow in Pipe

Providing a specific flow rate in a pipe is not a duty of the P&ID developer. It is, however, better to have a P&ID developer who also knows the fundamental of fluid flow to check the accuracy of the design while developing a P&ID.

The flow in a pipe cannot be set per our “wish” or the value we have in our mind or the “request” we mentioned in our mass balance table. The fluid flow is determined with a specific magnitude dictated by a pressure drive, which is a function of inlet pressure, outlet pressure, and pipe resistance. From a theoretical point of view, *all* flows can be applied in *all* pipes (with whatever pipe bore) as long as there is enough pressure drive. Therefore, a specific flow can be initiated in a piece of pipe as long as a specific pressure drive is applied to the pipe. Out of the three parameters that affect pressure drive, the inlet and outlet pressures are within the control of the designer. A P&ID developer can do a cursory check during the P&ID development

to make sure there is enough high pressure at a pipe inlet and enough low pressure at a pipe outlet to create the required flow rate.

Pipe resistance is generally beyond the control of the designer because it depends on the pipe material selected by the piping material specialist, the pipe diameter defined by economical measures, and the pipe length indicated in the plant plot plan.

6.5.2 Direction of Flow in Pipe

On P&IDs, each line representing a pipe usually has an arrowhead to show the direction of flow in the pipe based on the designed normal operation of the plant. However, these arrowheads can be confusing. As mentioned previously, flow always goes from a high-pressure (energy) point to a low-pressure (energy) point. Therefore, showing an arrowhead on a pipe does not guarantee that the flow always goes in the direction depicted on the P&ID. The direction of flow is based on the source and destination energies, irrespective of the designer's intention.

From here forward, the discussion will be based on pressure rather than energy because in the majority of cases to manipulate energy of a point, the pressure of that point must be manipulated.

For example, if there is a pipe connecting point A to point B (both at the same elevation) and the process design is such that the pressure at point A is higher than the pressure at point B, then the flow goes from point A to point B. However, if there is a control valve downstream of point A, then this control valve may drop the pressure at point A to a value lower than the pressure at point B. If this is the case, the flow will reverse back to point A rather than going to point B. This is an example of an action of a control valve. However, there could be several things that can change the pressure and eventually impact the flow direction. For example, a plugged filter can cause the pressure to drop and change the flow direction.

So, if there is a chance of a reverse flow in every single pipe, does a backflow prevention system like a check valve need to be installed? The answer is no. A backflow prevention system ensures the right flow direction. There are generally two cases where backflow prevention systems are placed: one is when the reverse flow damages a piece of equipment and the other is when the reverse flow is more probable.

If the direction of flow in a pipe is crucial and reverse flow is detrimental, then a backflow preventer should be used. For example, it is common to see check valves on the discharge side of centrifugal pumps and compressors because reverse flow damages the bearing of the shaft and also the electric motor.

How can a reverse flow be determined as more probable? The process engineer needs to review the P&ID. Some claim that there is no chance of a reverse flow in the primary (i.e. main and large bore) pipes. While it is true in a majority of cases (this is why there are fewer check valves on large-bore pipes), there are some exceptions. Meanwhile, others say there is a high probability of reverse flow in intermittent flow pipes.

Different backflow prevention systems will be discussed later in this chapter.

6.5.3 Providing Fluid with Enough Pressure at the Inlet

The pressure of a stream in a piece of pipe decreases gradually in the direction of flow. Therefore, if a specific pressure of P_B is planned at the destination point, there needs to be enough pressure at the point A, or P_A . P_A should definitely be higher than P_B . Point A may already have a pressure higher than point B, or it needs to be pressurized to a pressure higher than point B pressure. Therefore, the pressure in point B is out of human control. If the pressure in point B is lower than required, the pressure should be increased somewhere along the pipe and upstream of point B. If the pressure in point B is higher than required, the pressure should be decreased somewhere along the pipe and upstream of point B.

Increasing pressure can be done by pumps or compressors and decreasing pressure can be done by control valve or regulator. These items will be discussed in Chapters 7 and 10.

Every time the pressure of a stream is changed, the flow magnitude of stream will be changed, and vice versa. Such dependency always creates an issue in plant design because of preferred adjustment to pressure and flow rate independently due to their individual importance. Whenever pressure is decreased, the flow will be decreased, too, and designers have to make sure that the decreased flow rate is acceptable.

6.6 Piping Arrangements

Piping arrangements primarily can be classified as single-point border or multipoint border. Therefore, a pipe could be in different arrangements:

- Single-source, single-destination pipe (Figure 6.34a).
- Single-source, multiple-destination pipe (Figure 6.34b).
- Multiple-source, single-destination pipe (Figure 6.34c).
- Multiple-source, multiple-destination pipe (Figure 6.34d).

There are requirements that need to be addressed during the P&ID development of pipes. The easiest pipe arrangement is single-source, single-destination pipe.

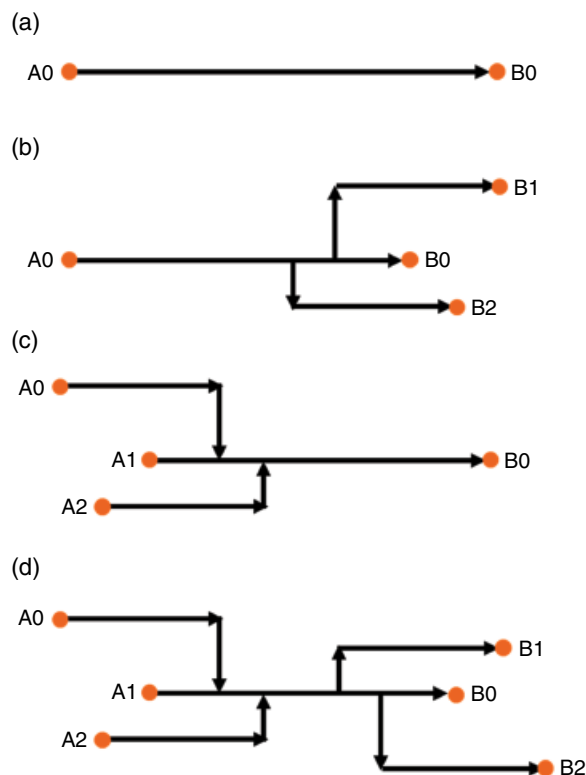


Figure 6.34 (a–d) Different piping arrangements.

In this arrangement, the first thing to be considered is making sure that the flow goes in the direction intended.

For all the cases of a single-source, multiple-destination pipe, multiple-source, single-destination pipe, and multiple-source, multiple-destination pipe having multiple points as source or destination is common. For simplicity, only the cases that have multiple destinations will be discussed. For these cases (as seen in Figure 6.35), the process requirement could be one of these goals: diverting or distributing the flow.

In diverting the flow, the goal is to transfer the flow to only one user at a time, while in distributing the flow, the goal is to direct the flow with different magnitudes to multiple users at the same time.

It means that in a multipoint source or destination in addition to making sure the flow is in the right direction, secondary goals should also be achieved.

The following sections explain how to satisfy the requirement of different pipe arrangements.

6.6.1 Backflow Prevention Systems [1]

When faced with the issue of reverse flow in pipes, the process design engineer usually uses a check valve to prevent the reverse flow. However, using a check valve is not the best method to prevent reverse flow. There are at

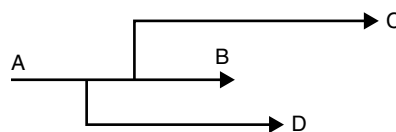


Figure 6.35 Multiple-destination pipe arrangement.

least four different techniques to prevent backflow, which will be discussed based on their level of effectiveness in preventing reverse flow.

#1 Air Gap: An air gap is the best way to prevent backflow. When an air gap is placed in a flow route there is absolutely no backflow in the system. However, this method has some disadvantages. The first is that an air gap is only applicable for liquid streams; they do not work on gas streams. The second is that there could be contamination of the liquid because the stream is exposed to the environment. Therefore, the environment should be clean, or the cost of cleaning up the contamination should be factored in. The other downside of an air gap is that if the weather is cold, there is a chance of freezing and system failure. However, if somehow all the disadvantages of an air gap are resolved, it provides one of the best backflow prevention techniques. An air gap could be placed in a pipe route or in a tank, and both are shown in Figure 6.36. It is common to see an air gap in potable water piping where there is a chance of backflow and contamination of portable water with other contaminated waters.

#5 Backflow-Preventing Device: Backflow-preventing device is a type of device that can be used in clean services and is common in water streams. In the casing of a backflow preventing device, there are two check valves in series and one pressure safety valve between two check valves in one casing. If a backflow-preventing device is used in a P&ID, it should be tagged as a specialty item (SP item) (Figure 6.37).

#2 Inverted U: An inverted U is another way to prevent reverse flow in a pipe. This method also works only on liquid streams. By creating a specific arrangement in a pipe, the backflow of a liquid stream can be prevented. This arrangement is a vertical upside down U. The height of this inverted U depends on the pressure of liquid flow when reversing back. Because of that, in some cases in which reverse flow pressure is high, the height of the upside down U will be beyond a practical value. In such cases, an inverted U cannot be used (Figure 6.38).

#3 Check Valve: Check valves are the most common way of preventing backflow. By placing a single check valve and reverse flow, either a liquid stream or gas stream will be prevented. However, a check valve always leaks when trying to prevent backflow. Generally it is assumed that the conventional swing check valve leaks about 10% of the flow. The other important parameter

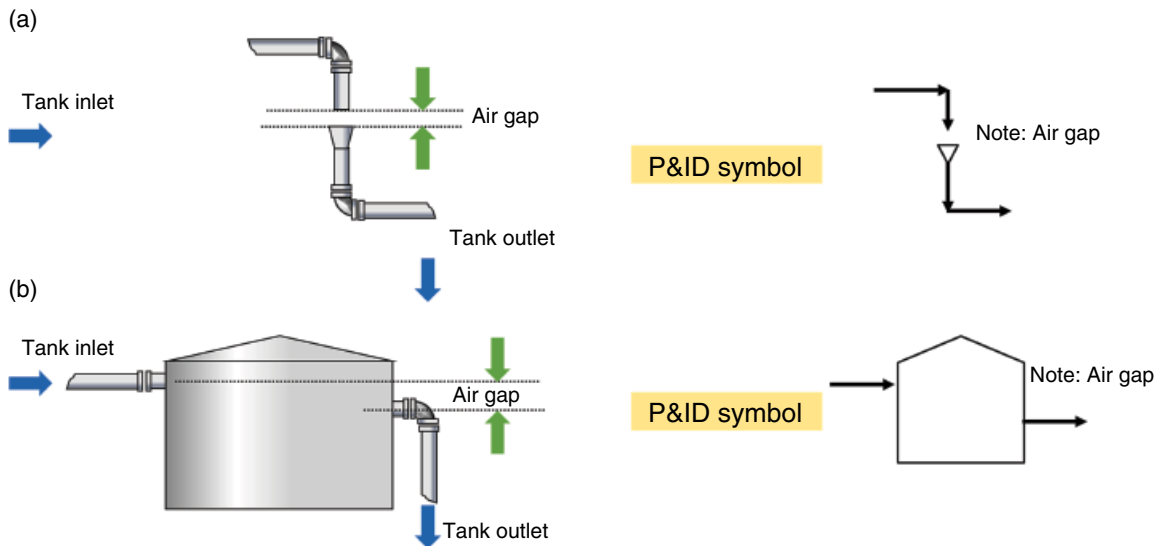


Figure 6.36 (a, b) Air gap.

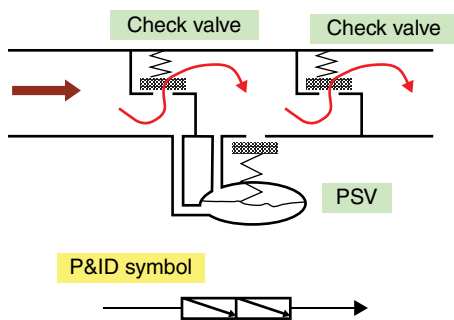


Figure 6.37 Backflow preventing device.

regarding a check valve is cracking pressure. Cracking pressure is the minimum pressure that a stream needs to have to be able to open up a check valve when a stream flows in a straight direction. It means that if the pressure of a flow is low enough, a check valve may not be a good choice to prevent backflow because the check valve may prevent even the straight flow. Because check valves generally leak at backflow, it is commonly said that check valves prevent liquid backflow, but they cannot prevent pressure penetration to upstream of the check valve. This concept is very important when deciding whether a

pressure safety valve should be placed when a check valve is involved (Figure 6.39).

Different types of check valves will be discussed in Chapter 7.

#4 Multiple Check Valves: There are some cases that two or three check valves can be placed in series to make sure the backflow is prevented and also to minimize the backflow leakage especially when the pressure of a stream is high. A conventional swing check valve may leak up to 10% of flow rate, so by placing two swing check valves in series, the backflow leakage can be decreased only 1%. If the pressure of the stream is less than 3000 kpag, use a single check valve. If the pressure is between 3000 and 7000 kpag, use double check valves. If more than that, a swing of three check valve is the most appropriate arrangement to prevent backflow. Some companies go further and indicate in their guidelines that whenever two or three check valves are used on a pipe, these should be dissimilar, meaning if one check valve is swing type, then the other one should be flow type. This guideline is meant to maximize reliability of the system because when there are two dissimilar check valves, there is less chance that both of check valves will jam at the same time (Figure 6.40).

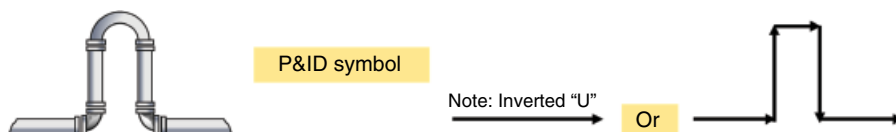


Figure 6.38 Inverted U for backflow prevention.



Figure 6.39 Check valve.

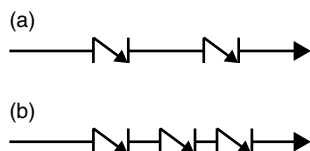


Figure 6.40 (a, b) Multiple check valves.

6.6.2 Diversion of Flow

The diverting of flow can easily be done by using several valves or a few multiway valves. One arrangement can be seen in Figure 6.41. In this arrangement, diverting valves work together to send the flow to destination C, B, or D. Each three-way valve could be replaced with two conventional (two-way) blocking valves. The multiway valves will be discussed in Chapter 7.

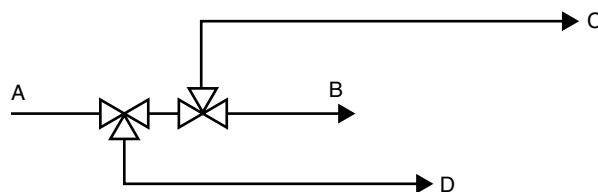


Figure 6.41 Diverting flow in a multiple-destination pipe route.

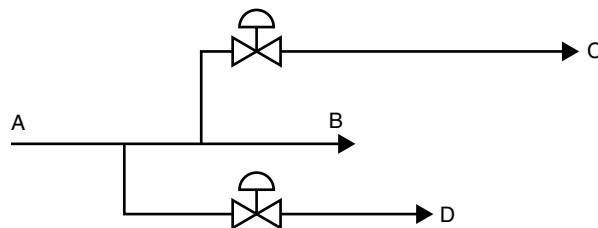


Figure 6.42 Distributing flow in multiple-destination pipe route by control valves.

6.6.3 Distribution of Flow

The distribution of flow is more difficult. From a purely theoretical point of view, it can be said that the solution is manipulating the branches' diameter and length to get to the desired flow rate. This solution, however, is not practical. The length of pipes is generally out of the control of the process engineer. The designer needs to needs to follow plant layout, which dictates the pipe lengths. Therefore, even though this solution may be financially attractive, it is not practical.

One magic bullet solution is to put control valves on all branches except one. This solution is shown in Figure 6.42. However, this solution can be expensive, especially when large bore pipes (larger than 12") are being used.

One branch can always be left without a control valve because when the flow rates of other branches are regulated by control valves, the flow rate of last branch will be regulated automatically.

It is important to consider that the branch without a control valve cannot be the route with maximum resistance. It means that when deciding on the branch without a control valve, that the branch does not get a minimum flow because the flow cannot be regulated in a pipe with constraints.

One alternative solution that *only* works when the intention is to distribute the flow *evenly* is to use symmetrical piping. The schematic in Figure 6.43 shows this solution.

The distribution of flow is not uncommon and wherever there is parallel operating equipment, this

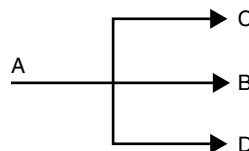


Figure 6.43 Distributing flow in multiple-destination pipe route through symmetrical piping.

requirement should be met. A note stating, "Symmetrical Piping," as in Figure 6.44 shows how this is used.

However, there are still some clients who prefer the control valve solution for applications like parallel heat exchangers and who are willing to pay the extra cost for that.

Placing a control valve to handle the flow distribution is known as *active distribution*, and when it is handled through symmetrical piping, it is known as *passive distribution*.

The best symmetrical piping arrangement for a precise flow distribution is the type in which the main pipe (before distribution) is perpendicular to the distribution header. However, in practice there is generally not enough room to implement such a concept. In practice, the main pipe could be connected to the distribution header after a run parallel to the distribution header and after an elbow. This arrangement generates a deviation from an evenly distribution goal.

Air coolers are a type of equipment that require good distribution because of their multiple inlets (and outlets). This will be discussed in more detail in Chapter 11.

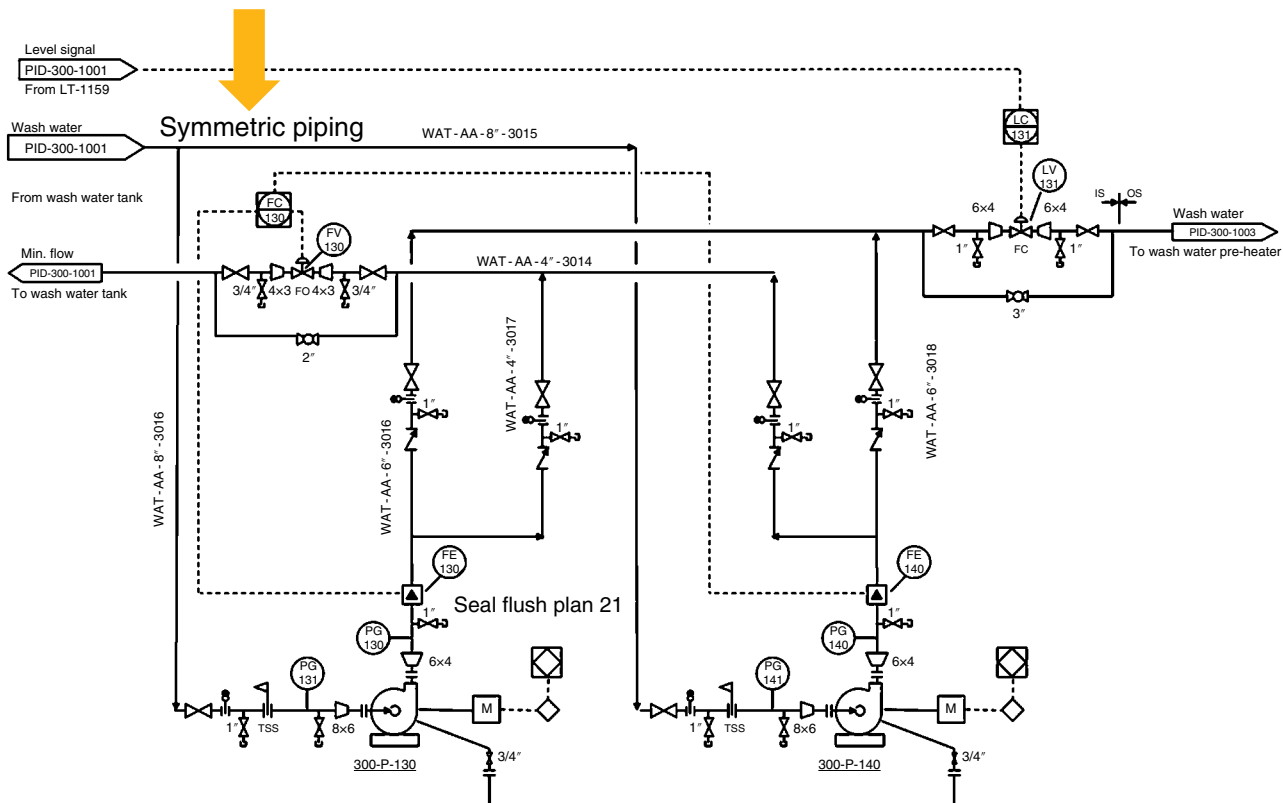


Figure 6.44 An example of symmetrical piping.

6.7 Pipe Route

Generally speaking, the real pipe route cannot be seen on the P&IDs. There are cases, however, that a special route should be considered for some routes based on engineering considerations. Such special pipe routes should be formally communicated with the Piping group (to develop proper pipe models and isometric drawings) to make sure they will be implemented in the plant during construction.

Because this route cannot be shown on the P&ID, such special requirements for pipe route should be captured on the P&IDs in the Notes.

A few of these special requirements are sloped, no liquid pocket, no gas pocket, free draining, vertical, horizontal, and minimum length or distance.

6.7.1 Slope

The slope on pipes can be important for horizontal pipes if a liquid or a two-phase flows with a liquid component flowing inside of them. There is no slope for vertical pipes, and slopes on gas flows are not important. Generally speaking, horizontal pipes are used without a slope for different reasons, including the difficulty in handling the pipe elevation and pipe supports. Therefore, by default no horizontal pipe will ever have a slope.



Figure 6.45 Slope symbol on pipes.

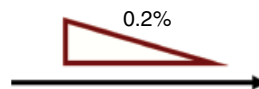


Figure 6.46 Slope symbol and slope magnitude on pipe.

There are two features that should be specified on a pipe symbol: the direction of slope and the slope magnitude.

The slope direction is the direction of flow (direct slope) or against the flow (reverse slope). The slope direction is shown on a pipe with a triangle symbol in Figure 6.45.

The other feature of a slope is its magnitude. From theoretical point of view, a slope can be specified by a percentage, by a fraction, or by the angle. However, angles (e.g. in degree) are not used to specify slope. A slope is specified either by a percentage or more practically by a fraction based on standard lengths of a pipe. The slope magnitude can be specified on the slope symbol on the P&ID as 0.5% or 2/12. Here, 12 feet is the standard length of pipes (Figure 6.46).

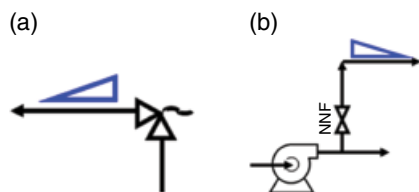


Figure 6.47 (a, b) Two examples of sloped pipe.

The direction of the slope can be decided during P&ID development, but slope magnitude should be calculated during the design.

The goal of sloping a pipe is to prevent stagnant liquid. In making a decision on the slope direction, the general rule is to put the slope toward the more tolerant resource.

For example, in Figure 6.47a, there is a need to slope the pipe in the outlet of pressure safety valves in liquid services or two-phase services (potentially or actually). In the majority of cases, the outlet flange of pressure safety valves dictates the horizontal pipe, and also the flow through the pressure safety valves are not continuous (the pressure safety valves will be opened during emergencies). Therefore, the outlet pipe should be sloped. It is not wise to have the remaining liquid stay at the outlet of the pressure safety valve, because the accumulated liquid at the outlet of pressure safety valve impacts its operation (refer to Chapter 12 for more details). Therefore, the slope of this pipe should be away of the pressure safety valve outlet.

Similar logic is used to put slope on a horizontal pipe, that is, normally no flow (NNF). This pipe goes to a vessel that is considered a more tolerant system for containing liquid; therefore, the slope is toward the vessel.

Another example is in steam piping. In steam distribution networks, there is always a chance of generating condensation in the pipe. Therefore, steam traps are installed at specific intervals to remove condensation from the steam flow, and the pipe is sloped toward each steam trap (Figure 6.48).

The concept of a steam trap will be discussed later in this chapter.

The last example is the pipe that directs released fluid from a pressure safety valve to a collection flare header

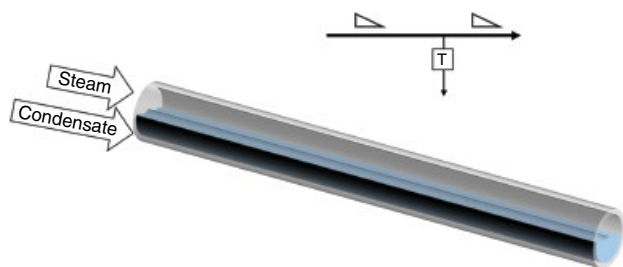


Figure 6.48 Example of sloped pipe in steam pipe.

(discussed more in Chapter 12). After collecting the release from different safety devices, the flare headers direct the liquid to the knockout drum before going to the flare. As the liquid knockout drum is placed to capture the liquid from the release gas, it is more tolerant to liquid and the main header is generally sloped toward the liquid knockout drum.

If the slope calculation is missed during the design phase of project, the process engineer responsible for P&ID development may decide to put a slope magnitude on pipes with triangles on them based on rule of thumbs. The minimum practical slope is about 0.08%, but the slope of the pipes can go up to 5%. The typical range of a slope is 0.5% and the typical range is between 0.2 to 1.0%. For hard-to-move or high-viscosity liquids, 2–3% is not rare. Underground pipes tend to be at the lower side at 0.2–0.5%. The reason is that if the slope of underground pipe is a large value, the pipe will be in deep ground at the destination point. Choosing a lower slope value is more important when the pipe is long or the underground water level is high (like near lakes).

6.7.2 No Liquid Pocket

A pipe fully carrying a liquid flow with the chance of excursion of gas or vapor can be a candidate for no liquid pocket, means “design in a way that liquid pockets naturally flow and exit the pipe route during the routine operation of the system”. This phrase directs the pipe modeler to design a pipe route wherein the gas flow cannot trap a pocket of liquid anywhere in the pipe route (Figure 6.49). To respond to this requirement, a piping modeler specifies a pipe route that is vertically downward or has a direct slope.

6.7.3 No Gas Pocket

A pipe fully carrying a gas or vapor flow with the chance of generation of liquid can be considered no gas/vapor pocket if the intention is to avoid a stagnant gas pocket in the pipe during the routine operation of the system. The term no gas/vapor pocket directs the pipe modeler to design a pipe route that the liquid flow cannot trap a pocket of gas anywhere in the pipe route (Figure 6.50). To respond to this requirement, a piping modeler specifies a pipe route that is vertically upward or has a reverse slope.

6.7.4 Free Draining (Self-Draining)

This note dictates the same requirements of no liquid pocket but during a system shutdown. This note actually means to do the piping in a way that no liquid remains in the pipe after a system shutdown.

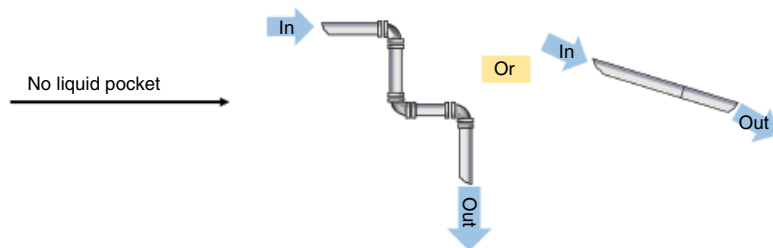


Figure 6.49 No liquid pocket piping.

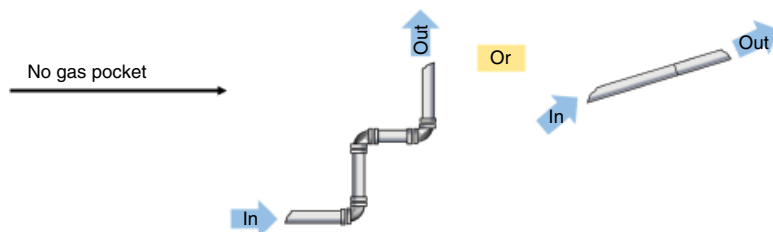


Figure 6.50 No gas pocket piping.

The piping route could be the same as no liquid pocket, but the sloped route is more preferable. If the “stairs type” route is selected, the horizontal pipes should at least have a slight direct slope.

6.7.5 Free Venting

This note also dictates the same requirements of a no gas pocket but during a system shutdown. This note actually means design the piping in a way that no process gas remains in the pipe after a system shutdown. This note is less common than free draining note.

6.7.6 Gravity Flow

The note gravity flow is abused in many cases. Figure 6.51 shows a flow driven by gravity. In such an arrangement, the fluid flows from tank A to tank B as long as the energy in tank A is higher than in tank B, or the liquid elevation is higher in tank A than in tank B. So, the pipe route does not change anything in this gravity flow. However if there is partial gravity flow, the pipe route should be a no liquid pocket.

6.7.7 Vertical or Horizontal Pipe

A note is required in cases where the vertical pipe is an outlet of pressure safety valves and connecting pipes to rotameters. It is common not to state this because the applicable cases are obvious to the majority of piping modelers. In some cases, instead of noting vertical pipe (or horizontal pipe), only a vertical (or horizontal) line is drawn on the P&ID, which is confusing because a P&ID does not show the route of a pipe.



Figure 6.51 Requirement of gravity flow and the pipe routes.

6.7.8 Straight Piping

This is a common note when dealing with flow meters. Flow meters generally have a required straight pipe in their upstream and downstream. Such a straight portion of pipe should be considered during the piping modeling. However, in the majority of cases, such note is not stated as the piping modelers are well aware of this requirement.

6.7.9 Minimum or Maximum Length or Distance

An example of this was discussed already in regard to the pipe spec border extension. The other example is minimizing the dead end portions on piping in a slurry or perishable fluid (mainly liquid) systems (Figure 6.52).

The other example is the distance between the primary control element (sensor) and the manual throttling valve in the control valve stations. As will be discussed in Chapter 7, during control valve off operation, an operator needs to watch the sensor and throttle the manual valve, hence bypassing the control valve. Then, the sensor and the manual throttling valve should be close to each other. Such a requirement could be detailed with a note about the minimum length for the pipe between them or a “readable from valve” note near the sensor (Figure 6.53).

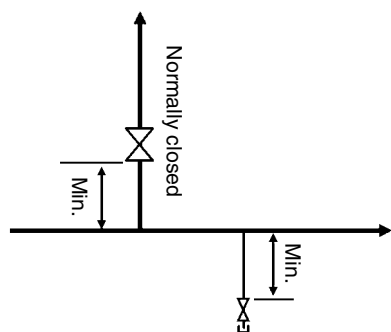


Figure 6.52 Min. length of a note to eliminate dead end.

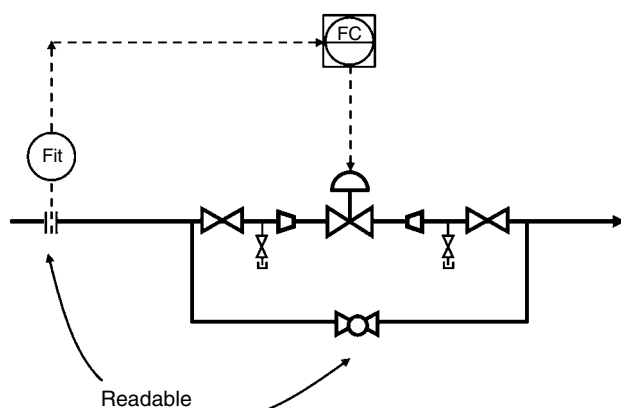


Figure 6.53 Min. length of a note for control valve stations.

6.7.10 Other Special Pipe Routes

A U-shape on a P&ID does not actually represent the installing of a U-shaped pipe. If a piece of pipe with special routing will be provided by a vendor, the data sheet will show the details of the pipe. However, if the special pipe route will be fabricated by Piping staff in field, it should be mentioned in the P&IDs through a note. It is common to ignore the note and show the real pipe routing on the P&ID by relying on the knowledge of the pipe modeler.

6.8 Piping Movement

A piping circuit can be moved slightly during operation because of different reasons. One cause of a moving pipe circuit is thermal expansion and the other is equipment movement.

Because of the high temperature of service fluid or temperature variation of ambient air, thermal expansion can expand the piping circuit. Such expansion can lead to breaks in the pipes if there are no provisions to handle that.

Equipment movement is another reason for pipe movement. If a piece of equipment connected to a pipe is

moving, then pipe will move. Again, if this pipe movement is not planned for and stopped in time, the pipe will be break. There are two main types of equipment that may cause movement: rotary machine vibration and settlement of big footprint equipment.

Equipment like pumps, compressors, and centrifuges may continuously vibrate. This vibration will transfer to all connected pipes if it is not mitigated. However, with today's technology and placement of equipment on the skids, the vibration of equipment is rarely a problem.

The well-known example of a big footprint equipment are large tanks. Tanks with a large diameter are considered big footprint equipment. Several years after tank fabrication, one side of the tank may settle higher or lower than the other side of the tank. This may lead to a marginally tilted tank and the connected pipes are displaced and will eventually break. Tanks settle due to the nonhomogenous nature of the soil and the imperfect foundation of the tank. Such settlement is not uncommon and can often be seen in buildings throughout the world. A few years after a new house is built, some cracks can appear in different locations of the house, and this is a result of the settlement of the house.

There are two ways that pipe movement can be mitigated:

- 1) Shifting movement: By placing expansion loops on a pipe circuit, the unwelcomed movement of pipe is transferred to the elbows, which can handle those movements better than a straight pipe. Expansion loops could be in the form of horizontal U-shapes (Figure 6.54).
- 2) Isolation of equipment: By isolation of the pipe circuit from the moving equipment, the movement will not be transferred to the pipe circuit. Flexible connections are used to isolate a piece of equipment. Flexible connections have symbols, and they are shown in Figure 6.55 and should be tagged as a SP item in the

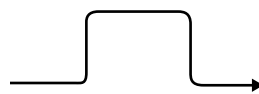


Figure 6.54 Expansion loop in pipes.

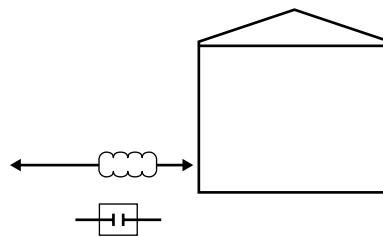


Figure 6.55 Flexible connection on a large bore pipe connected to a tank.

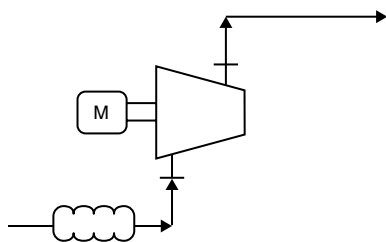


Figure 6.56 Flexible connection on the inlet of centrifugal compressor.

P&IDs. For example, some company guidelines ask for connecting pipes larger than 4 or 6 in. through flexible connections when they are connected to tanks to absorb the settlement of the tank. Small bore pipes are waived from this guideline because they can handle a few tank settlement by creating a sagging in the pipe.

Figure 6.56 shows a flexible connection on the inlet of a centrifugal compressor, although it is not common these days.

Flexible connections used to have a bad reputation regarding leakage. But now there are better flexible joints in the market.

6.9 Dealing with Unwanted Two-Phase Flow in Pipes

The design and implementation of systems in two-phase flows are more difficult than single-flow pipes. There are, however, cases in which a two-phase flow is inevitable.

When the flow is intended to be a single flow, but then it turns out to be a two-phase flow, the piping design is based on a single phase, and the two-phase flow should be eliminated. There are three types of two-phase flows: liquid–gas, gas–liquid, and solid–liquid.

6.9.1 Liquid–Gas Two-Phase Flow

In a liquid–gas two-phase flow, there is a chance of liquid droplets in the main stream of gas or vapor. The problem arises when transferring a gas or vapor because a liquid can be generated and that is problematic.

Such unwanted two-phase flows may happen at different times. One is when gas comes off of a liquid surface, like in liquid–gas separators. The other case is when transferring hot vapors, like steam.

The first step in dealing with this problem is to prevent the creation of a two-phase flow. For example, when transferring a wet gas, heat trace (dashed line beside the

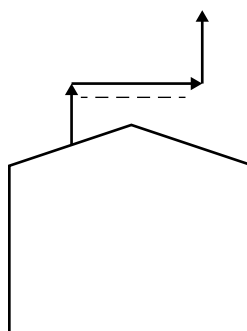


Figure 6.57 Heat tracing to prevent the generation of condensation.

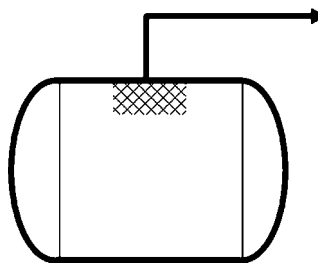


Figure 6.58 Demister to prevent carrying over of liquid droplet.



Figure 6.59 Steam trap action.

main line) may be used. This solution can be seen in Figure 6.57.

The next method is to remove the generated liquid phase from the gas phase as soon as possible before the creation of a slug of liquids. One example is using a demister in a gas–liquid separator vessel as shown in Figure 6.58. For gas streams that come off of a liquid surface, there is always the chance of carrying liquid droplets over into the gas stream.

The other example is using steam trap in steam distribution piping networks. Steam traps remove water condensation from the steam (Figure 6.59).

In Figure 6.60, a steam trap is shown as a square with letter T at the middle. Steam traps should be installed at predetermined distances on steam transfer pipes, and the pipes should be sloped toward the steam traps. Failure to install a condensation removal system in steam pipes may lead to steam hammering, which may break the pipes.

There are other symbols can be used on P&IDs for steam traps if the intention is to use the exact type of steam trap (Figure 6.61).

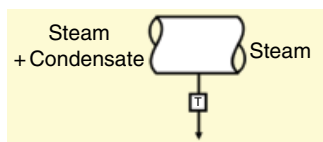


Figure 6.60 Steam trap in steam pipes.

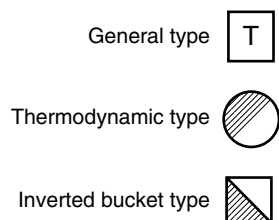


Figure 6.61 P&ID symbols for different types of steam traps.

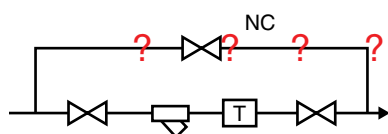


Figure 6.62 Bypass for steam traps.

In these applications, steam traps are installed on vertical short pieces of trim pipe known as a steam trap installation on “drip legs.” Steam traps, however, can be installed on main pipes. The arrangement of a steam trap in such an application is more complicated than the steam trap on a drip leg.

Steam traps need a strainer on the upstream to protect them from jamming (will be discussed in more detail in the next section). Drip legs already have a mud column on their bottom and the steam traps installed on them do not need a strainer.

They also may need an isolation valve (Chapter 7) for steam trap inspection, cleaning, or maintenance. However, adding a bypass pipe loop is always debatable and will need to be considered fully (Figure 6.62).

In the absence of a steam trap, the isolation valves are closed and the bypass valve is open, and then the two-phase flow bypasses the steam trap and goes downstream. It should be ensured that the piping and equipment in downstream does not damage the pipes or equipment. In many cases, companies forbid providing a bypass for steam traps.

Steam traps are a compact version of a two-phase separator, which does not need a control system and it only works on condensate two-phase streams. This is shown in Figure 6.63.

Usage of slug catchers can be explained with the same logic. Slug catchers are a common piece of equipment in natural gas pipelines. Within the natural gas transfer

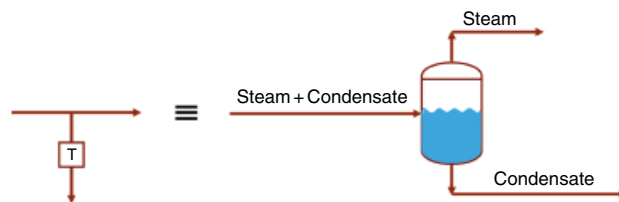


Figure 6.63 Steam trap function concept.

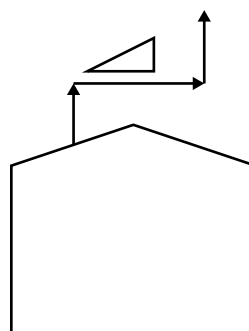


Figure 6.64 Sloped pipe to direct the liquid phase to more tolerant system.

through the pipeline, some less volatile components of the natural gas turn into liquid. Therefore, at some points, the pipeline may have a two-phase flow. Although the natural gas pipeline may be designed to handle the two-phase flow of gas–liquid, this not the case for the equipment downstream of a pipeline. Therefore, there could be a slug catcher at the end of natural gas pipeline to remove the liquid portion of the stream from the gas portion.

The next solution is directing the created liquid phase to more tolerant system. A liquid phase together with gas phase is not good for pipes, but the liquid phase can be redirected to a liquid tank, which is a system that can handle (and is designed to contain) liquid. In Figure 6.64, the pipe sloped toward the tank to direct the generated liquid back to the tank.

The other solution is to ignore it. With this solution whenever a branch is used from a steam header, the connecting point is put on the top of the header to make sure the generated condensation is not drawn from the header and the user only receives the steam. The example is when branches hook up on a flare header or branches from a bit wet plant air. Generally, the connection point between the branches of the header is on the top of the header. Only a note can capture this point on the P&ID.

In such cases, using some fitting in the pipe route should be avoided. For example, concentric reducers are harmful in two-phase flow streams. The concentric reducers may change the two-phase flow from a less

detrimental regime of dispersed to a more detrimental regime of slug flow. If placing a reducer is needed on a potential two-phase liquid–gas flow, it should be a flat-on-bottom (FOB) one.

6.9.2 Gas–Liquid Two-Phase Flow

Gas–liquid two-phase flows may occur in liquid flows with a chance of generating or ingressing gas bubbles in them. This issue is much easier to solve than the previous case because the gas bubbles have a natural tendency to go up. Therefore, they can be removed from the liquid flow easily. Gas bubbles go up and accumulate in the corner areas of piping. A gas release system can be placed in those high point areas to remove the collected gas bubbles. If the accumulation of gas bubbles is not very much, the gas release system can be a simple vent valve. If the accumulation of gas is higher, a gas release valve or even an automatic time-driven valve can be placed. Installing air release valves are common in water pipes (Figure 6.65).

In these cases, using fittings in the pipe route should be avoided; for example, concentric reducers are harmful in two-phase flow streams. Concentric reducers may change a two-phase flow from a less harmful flow to a more hazardous slug flow. If placing a reducer is needed on a potential two-phase gas–liquid flow, it should be a flat-on-top (FOT) one.

Such unwanted two-phase flows may happen mainly when the liquid is cold and is in contact with a gas (or air). The colder liquid can dissolve more gases, and when the stream goes through pipes, the pressure gradually drops and the dissolved gases turn into bubbles. Such issue is more problematic in pipelines with long pipe lengths. One of the cases of gas–liquid flow could be in suction side pumps when trying to pump a liquid with dissolved gases. The other case in which an unwanted gas–liquid flow may be created is when liquid flow with dissolved gases goes through filters or packed columns. In such situations, each streamline has a random route; the longer streamlines experience more pressure drops and the dissolved gas turn into bubbles.

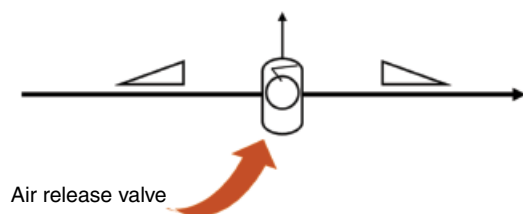


Figure 6.65 Air release valve.

6.9.3 Solid–Liquid Two-Phase Flow

The solutions to this issue are similar to gas–liquid solutions. One is getting over the existence of solid phase and taking the branch from the top of the header, wherever there is a need for reducer using an FOB type of reducer on the pipe or making the pipe vertical.

Another solution is removing the solids through a strainer. This strainer is not a process step, but only for equipment protection. The bad thing about strainers is that they are often overlooked, and no one removes and cleans them when they are full.

Removing a suspended solid by installing a strainer is unnecessary unless the downstream piece of equipment has a tight clearances that may plugged and jammed quickly (e.g. centrifugal pumps, PD pumps, steam traps, some types of flow meters like turbine flow meters.)

Such unwanted two-phase flows may happen when dealing with dirty services. However, the flow during the commissioning of a process plant is generally dirty. Commissioning is the first start-up after the construction of a plant. Therefore, the fluids in pipes carry debris, leaves, sandwich wrappers, or used welding rods. Such large suspended solids can be harmful to the equipment. Thus, in some cases, even when the flow is clean during the normal operation of the plant, it could be dirty during the commissioning.

The strainers that are placed for the purpose of protecting the downstream during normal operation are called permanent strainers, and the strainers for commissioning are known as temporary strainers. Temporary strainers are supposed to be removed after commissioning.

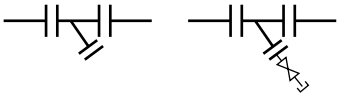
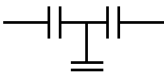


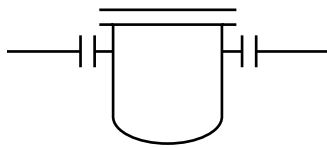
Table 6.7 shows the application of different strainers for temporary and permanent applications for small and large bore pipes.

It is important to note that strainers are not installed to protect items that are already built, but rather to deal with suspended solids. For example, a strainer upstream of a slurry pump can be questionable.

6.10 Tubes

As was mentioned previously, tubes are pipes with sizes less than 2". A general rule of thumb is that a P&ID lacks callouts for tubes. However, there are cases where some tubes may be shown, but they are still not tagged. A well-known example is instrument air tubing. Instrument air generally is distributed through pipes, and they are shown on the P&IDs. However, near the instrument air users (like control valves), the instrument air pipe is connected to a manifold that is used as starting point for several instrument tubes to the instruments. These tubes

Table 6.7 Applications of different types of strainers for protection.

Pipe size	Temporary	Permanent
NPS < 3"	Y-type strainer 	T-type strainer 
NPS > 6"	Cone-type strainer  Or 	Bucket-type strainer 

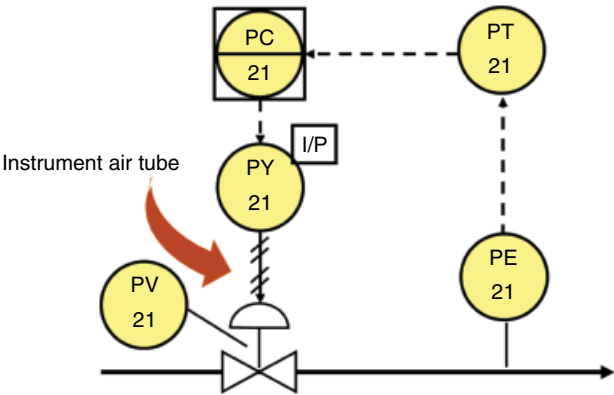


Figure 6.66 Instrument air tube inside of control loops.

are generally not shown unless inside of a control loop (Figure 6.66).

A “footprint” of tubes may be seen in hot fluid tracers for piping circuits. When a pipe is steam traced or hot glycol traced, it can be recognized from the pipe tag or from a little symbol on the pipe. The symbol generally has a short line that represents the heat-tracing tubes. In other cases, the heat-tracing tubes may be shown as a parallel dashed line beside the process pipe. Later, that dashed line will represent an electrical signal, but this dashed line representing tracers will not be confused with the electrical signals because the tracers always are shown beside the pipes following their routes (Figure 6.67).

The tubing that goes inside of some instruments, like sensors, is also shown. These narrow tubes are called the capillary and are shown in Figure 6.68. More information about this will be provided in Chapter 13.

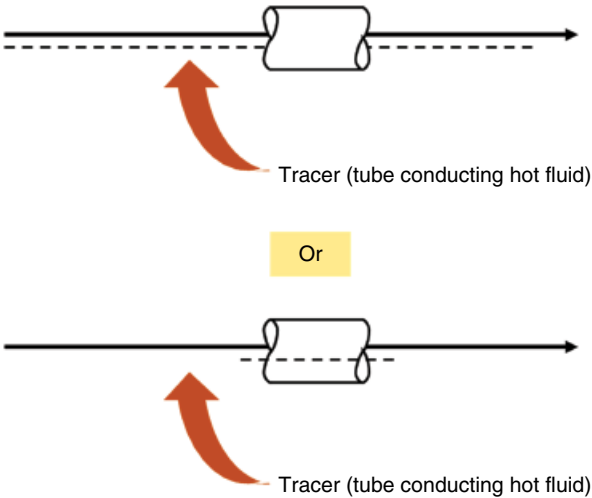


Figure 6.67 Fluid heat traced pipe.

The other situation that a tube may be used is when discussing a circular cross section fluid conductors inside heat transfer equipment. Heat exchanger tubes and fire-heated tubes are example of these applications.

6.11 Double-Wall Pipes

Double-wall pipes are available for different reasons. One application of double-wall pipes are when jacketing is needed. For example, in heat jacking, a heating media, like steam, flows into the annular space, whereas the main stream flows into the internal pipe of double-wall

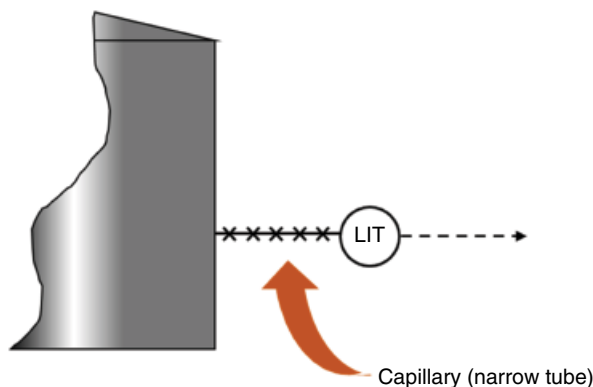


Figure 6.68 Tubes or capillaries for instruments.

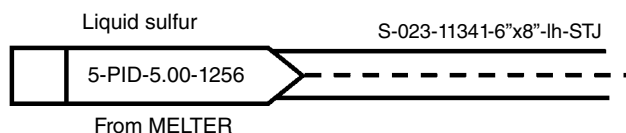


Figure 6.69 Tagging a double-wall pipe.

pipe. As a rule of thumb, when the temperature difference between the transferring fluid and the ambient temperature is more than 100–150°C, insulation is not effective and heat jacketing (e.g. steam jacketing) should be used. One example of using double-wall pipe for this purpose is transferring molten sulfur.

The other application of double-wall pipes are when the intention is to provide a secondary containment for pipes. Secondary containment will be discussed in Chapter 9, but in a nutshell, it is an additional enclosing wall around the process fluid. One example of using double-wall pipe for this purpose is transferring of toxic fluids.

The last application mentioned here is using a double-wall pipe, whereas the annular space kept under vacuum works as an insulation layer around the main internal pipe. This application is more common to provide cold insulation rather than hot insulation. One example of using double-wall pipe for this purpose is cryogenic ethane. Double-wall pipes can be tagged separate (each pipe, inner and outer separately) or within a single tag (Figure 6.69).

6.12 Pipes for Special Arrangements

Pipes can be placed to provide different features.

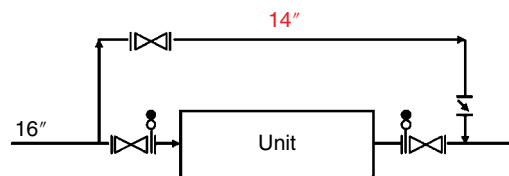


Figure 6.70 Unit bypass pipe.

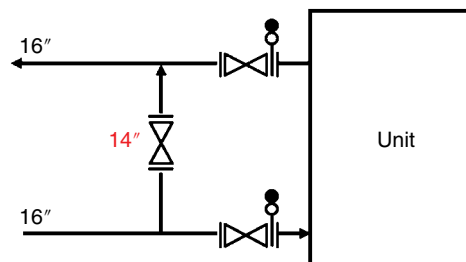


Figure 6.71 Recirculation bypass pipe.

6.12.1 Piping for Bypassing

One of the available options when a piece of equipment is out of operation is providing a bypass for it (Figure 6.70). The bypass pipe could be smaller than the main pipe size (e.g. one size smaller) because there is no pressure drop creating an item on it.

The bypass pipe should have at least one blocking valve on it near its beginning. If the bypass pipe is long or there is a fear of dead leg, it may need another blocking valve near its end.

By providing a bypass pipe, the flow automatically goes through it because the pipe directs the flow to downstream, which has a lower pressure. However, if there is any chance of using the bypass pipe during the normal operation of the unit, it may need to be equipped with a check valve to prevent backflow.

Such bypass pipe is needed for closed recirculating systems (Figure 6.71). This is to give flexibility to the fluid to the system in recirculating mode or bypassing the unit.

6.12.2 Piping for Recirculation

Recirculation can be needed for different reasons. It may be part of the normal operation of a unit for recycling and increasing the efficiency of the unit. Recirculation may be needed for starting up or cleaning a unit. If recirculation is needed during normal operation, the unit most likely needs to be equipped with a recirculating pump or a recirculating compressor (Figure 6.72).

6.12.3 Piping for Units in Series

When two (or more) units are in series, the last one generally works as a trimmer. It is common to see a bypass

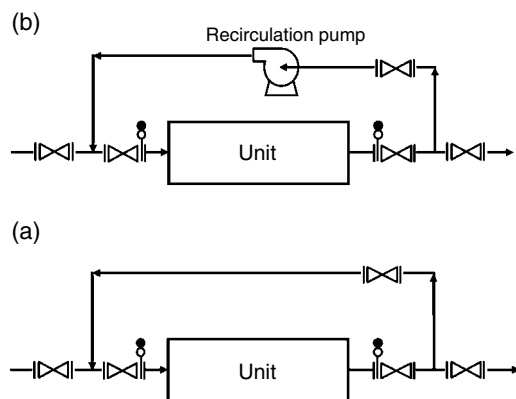


Figure 6.72 (a, b) Unit recirculation pipe.

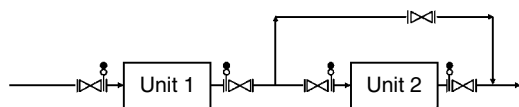


Figure 6.73 Series units pipes.

pipe around the trimmer unit. This bypass pipe can be used during the start-up of the plant when there is no real need to bring all the units in operation at once, and the trimmer unit can be bypassed (Figure 6.73).

6.12.4 Piping for Units in Parallel

Units can be placed in parallel, and their piping is a bit tricky. Here we want to focus on the pipe sizes in such arrangement.

It is obvious that if there are two similar equipment in parallel and one of them is the operating piece of equipment and the other, a spare one (it means $2 \times 100\%$ sparing philosophy), the pipe size does not change after splitting. This concept is shown in Figure 6.74.

But if both pumps are operating (it means $2 \times 50\%$ sparing philosophy), the flow splitting on the inlet side of the pipes in a way that each pump will receive half of the flow. Here the rule of thumb says that the size of each branch is $\sqrt{2}$ of the size of main header, as shown in Figure 6.75. Based on this, when a pipe flow is branched to three even flows, the size of each branch would be $\sqrt{3}$ of the size of main header.

6.12.5 Piping for Pressure Equalization

The pipe for pressure transfer or pressure equalization can be much smaller than the main pipe size (i.e. two or three size smaller).

The pipe for pressure equalization can be connected to two sides of blocking valves for ease of opening or between two tanks to allow initiating the liquid flow between them.

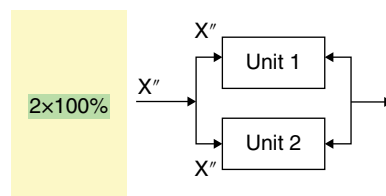


Figure 6.74 Pipe sizes of parallel units, a spare one operating.

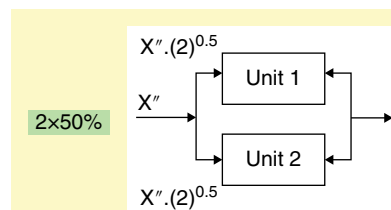


Figure 6.75 Pipe sizes of parallel units, both operating.

6.13 Pipe Size Rule of Thumbs

It is not easy to tell from a P&ID whether the pipe size is correct. However, when two or more pipes are connected to each other, it is easier to check the accuracy of pipe sizing.

Below are some cases through which the pipe size can be checked:

- 1) When two (or more) pipes are merging together, the resultant pipe may have a larger size. Here the word *may* is used because there are some cases that this rule is not valid. For example, when a 2" pipe is merged to a 20" pipe, the size of pipe after connection of the 2" pipe is less likely to be changed to a larger size, like 22".
- 2) When a pipe is split into two or more branches, the size of branches may be smaller than the main pipe.

6.14 Pipe Appurtenances

Pipe appurtenances are mainly classified into three main groups of valves: fittings, process items, and non-process items.

Valves are piping components that actively affect flows. *Active* means they should have a movable part. A gate valve has a moving stem. A check valve has a moving flap.

Fittings are piping appurtenances that passively affect flows. Examples of fittings are elbows and reducers.

Process items are piping appurtenances that are installed on piping and change some process feature of the flow. Examples are strainers and silencers. Process items generally tagged in P&IDs as SP item.

Nonprocess items are the items that do nothing to flowing fluids. Different types of pipe supports and pipe hangers can be classified in this group. Because they are not generally shown on the P&ID, they are not discussed here. The only nonprocess item that we discuss here is an insulation joint.

Fittings and special items are discussed in this chapter, and valves will be discussed in Chapter 7.

6.14.1 Pipe Fittings

Fittings are generally not shown on P&IDs except for four different groups of fittings: reducers and enlargers, three-way connections, end-of-pipe systems, and process flanges. If there should be specific fitting on a pipe and it cannot be shown on the P&ID, it can be covered in the Note area of the P&ID sheet.

6.14.1.1 Pipe Direction Change

No pipe direction change can be seen on P&ID. However, there could be some cases in which the process dictates using or not using a specific type of pipe direction change fitting. In such cases, the requirement is put in a note about the pipe route on the P&ID.

The most common pipe direction change is a 90-degree redirection fitting; however, a 45-degree fitting is available, too. A 90-degree elbow can be standard, short, or long.

The piping designer always chooses the standard elbow except in specific cases. When the space is tight, a short radius rather than standard radius is used. The long radius is placed only if the P&ID developer requests it. The process designer may ask for long radius for specific cases like slurry piping. The standard radius elbows are quickly abraded in some slurry piping, and it is wise to use long radius.

6.14.1.2 Reducers (Enlargers)

The goal in using reducers or enlargers is to increase or decrease the pipe size (pipe diameter). There are at least four situations that reducers or enlargers are used.

The first situation is in the case of tying-in one or more pipes to another pipe and branching a pipe to more pipes. In these situations, a reducer or enlarger needs to be used (depends on the case) to keep the same velocity in the pipe after adding or deducting fluid flow from the pipe.

If a pipe needs to be branched off to other pipes, it means that a portion of the flow will be directed from the main pipe (header) to other pipes, leading to the decrease in the flow in the main pipe. If the diameter of the pipe is not decreased, the velocity of fluid in the pipe will be decreased, which is not a good idea. Therefore, the reducer is used to adjust the pipe size with the new reduced flow rate in the main pipe or header. The reducer

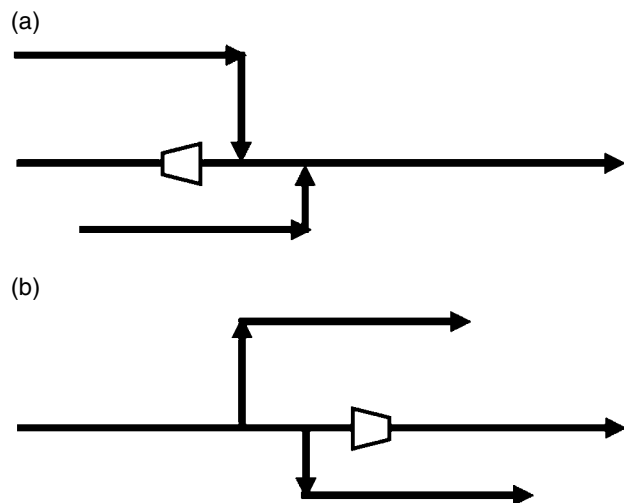


Figure 6.76 (a, b) Tying-in and branching-off piping arrangement.

should always be placed after pipe branches and not before them. This can be seen in Figure 6.76a.

In a tying-in situation, there are one or more pipes that merge in the main pipe or header. In this situation, the flow rate of the main header is increased after the addition of flow through the tying-in pipes. Again in this case, an enlarger is used. The important point here is that the enlarger should be placed before the tying-in pipes not after them, as in Figure 6.76b.

The second situation that a reducer or enlarger is used is when a pipe is connected to a piece of equipment or instrument. There are some cases that the equipment flanges do not match the pipe flanges; therefore, a reducer or enlarger needs to be able to connect pipe to a piece of equipment, such as connecting suction piping to a centrifugal pump or connecting discharge piping to the centrifugal pump. Other examples are piping connections to inlets or outlet of pressure safety valves and the pipes upstream or downstream of some control valves. These examples are shown in Figure 6.77.

We can size a pipe to make sure it matches the equipment or instrument flanges, and by doing this, the requirement of reducer or enlarger may be waived. However, this is not always feasible. Equipment and instruments are sized and manufactured based on the best equipment or instrument. The sizing criteria for sizing a piece of equipment or instrument are not

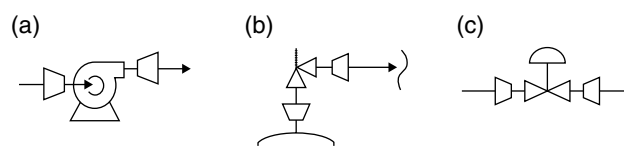


Figure 6.77 (a–c) Examples of reducer applications to match the manufacturers.

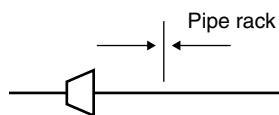


Figure 6.78 Reducer applications for pipe sturdiness.

necessarily the same sizing with those for pipe sizing. This is why pipe sizes do not always match the equipment or instrument flanges.

The third situation that we need to put enlarger or reducer is to support specific mechanical strength for a piece of pipe. For example, if a pipe should go over a pipe rack, it should be a minimum of 2" in size, because, as a rule of thumb, if a pipe is smaller than 2", it will sag when it goes over the pipe rack and may need additional support to prevent this sagging. However, such additional support is not always available. There are some cases that a process engineer sizes a pipe and later uses a size less than 2", for example, 1 1/2" pipe that goes from point A to point B. However, if this pipe goes over the pipe rack, an enlarger may possibly need to be put on the pipe to increase its size from 1 1/2" to 2" just before going over the pipe rack. And after coming down from a pipe rack, a reducer may be used to change the pipe size back to what the process engineer designed. This example shows a case where both reducer and enlarger were used to support the specific mechanical strength of a piece of pipe (Figure 6.78).

The fourth situation is using a reducer or an enlarger for process purposes. When a pipe is narrower, the velocity is higher, and vice versa. There are some cases in which pipes connected to equipment and the pipe is enlarged before connecting to the equipment possibly to shorten the fluid jet length inside the equipment if the long jet disturbs the process inside. In other cases, enlarging a pipe before connecting it to a tank to reduce or eliminate the generated electrostatic charge is especially important when dealing with flammable liquids. If the inlet nozzle of a tank is narrow, an electrostatic charge may generated and cause fire.

However, the process engineer is not always neutral about using reducer or enlarger. Using an enlarger, the velocity of fluid goes below the acceptable value, making it unusable. In slurry piping, if the fluid flow velocity goes below a specific value, sedimentation occurs and that may lead to the plugging of the pipe.

The symbol for reducers and enlargers are trapezoid. The sizes of reducers are mentioned below each reducer or enlarger symbol in a P&ID. The way the size is shown is to mention the larger side of reducer or enlarger and a multiplication sign (×) and the smaller size of the enlarger or reducer. The examples are shown in Figure 6.79.

However, during the P&ID development, sometimes a process engineer asks, What size of reducers or enlargers

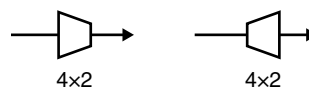


Figure 6.79 Reducer or enlarger size on P&ID.

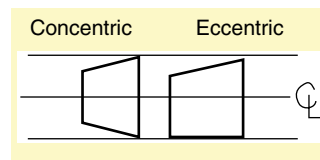


Figure 6.80 Two types of reducers.

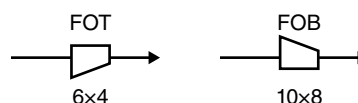


Figure 6.81 Showing eccentric reducers on P&IDs

are available? Can a reducer be used that reduces the pipe size from 20" to 1"? This is best answered by referring to piping documents; however, it is a good idea to have a rule of thumb to have some idea about the available reducer and enlargers. Generally speaking, the minimum size that a reducer can decrease the size of a pipe is half of that pipe size. It means with a 20" pipe, the reducer, which can be changed more severely, this pipe size is 20/2 or 10", which is 20 multiplied to 10 reducers. It means for this case, the available reducers are: 20 × 18, 20 × 16, 20 × 14, 20 × 12, and 20 × 10".

Reducers and enlargers have two main types: concentric and eccentric (Figure 6.80). Concentric reducers are less expensive than the more common reducers because it is a symmetrical piece and fabricating symmetrical things is cheaper than fabricating nonsymmetrical things. Therefore, concentric reducers or enlargers are used unless instructed otherwise.

The difference between concentric and eccentric reducers is important only in horizontal pipes. When using eccentric reducers (on horizontal pipes), it should be cleared if the flat portion is at the bottom or at the top of the fitting. The position of flat portion of reducers or enlargers is shown through the symbol or by acronyms of FOT or FOB near the symbol or both per the company guidelines (Figure 6.81).

What should be done when the less expensive concentric reducer (or enlarger) cannot be used and the more expensive eccentric ones have to be used instead?

One main reason is the cost of adjusting pipe supports for changing the pipe size because of concentric reducer/enlarger. When/if we use concentric reducer or enlarger, the pipe supports and hangers should be adjusted

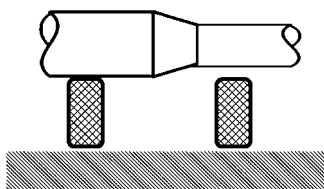


Figure 6.82 Need for eccentric reducer to use identical pipe supports.

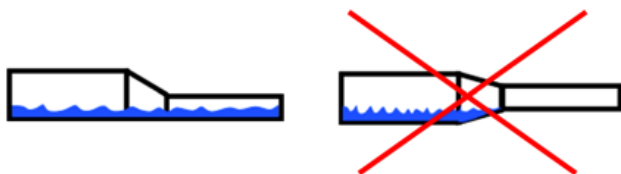


Figure 6.83 Using FOB eccentric reducer or enlarger to satisfy full draining of liquid pipes.

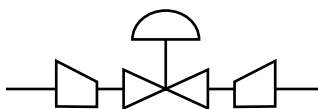


Figure 6.84 Using FOB eccentric reducer for control valve.

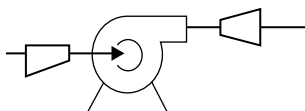


Figure 6.85 Using FOT eccentric reducer in suction of centrifugal pump.

accordingly. If for whatever reason there is financial constraint for designing and also buying different pipe supports, eccentric reducer or enlarger with FOB can be used instead (Figure 6.82).

Another reason that may dictate using eccentric reducer or enlarger is specific for liquid transferring pipes. An example is facilitating the draining operation of a pipe circuit. For liquid transferring pipes, the reducer or enlarger that is installed FOB can solve the problem. An example of this application is for the reducers of control valves (Figure 6.84).

The full draining is not always important. If the pipe is a small bore (possibly less than 4") or the liquid is innocent, it is possible that there is no need to provide full draining capability such as also in pumping a liquid with dissolved gases as it was discussed previously (Figure 6.85).

The discharge of the centrifugal pump possibly does not need the eccentric FOT enlarger because it is most likely a vertical pipe.

6.14.1.3 Three-Way Connections

In Section 6.9, a multipoint source and multipoint destination pipe arrangements are discussed.

When there is a pipe arrangement with multiple sources or multiple destinations, the concept of tying-in and branching off comes to the picture. Both of these concepts boil down to the question: How is a pipe connected to the middle of another pipe? Generally the response would be a requirement for a three-way connection. But there are several types of three-way connections. To know which type of three-way connection should be used, a P&ID developer needs to consult the branch table available in the piping material spec table of the project. There are different fittings available to do that including tee's, reduced-tee's, and different types of O-lets (e.g. Weldolets, Thredolet).

However, Table 6.8 can be used in deciding on the type of three-way connection to be used, and it also shows the representation of different three-way connections on P&IDs.

As can be seen from the table, the type of intersection from P&ID cannot be figured out. The three-way connection could be a tee, reduced tee, or O-let. There is no specific symbol to recognize a tee from a reduced tee or from an O-let. Therefore, a P&ID will not show whether a three-way connection is a tee, reduced tee, or O-let. If someone wants to see if the intersection is tee, reduced tee, or O-let, the piping spec must be consulted.

6.14.1.4 Pipe Connections

The pipe manufacturers do not fabricate pipes in infinite lengths. During the construction, the pipes should be connected to each other, end to end, to make a suitable length of a pipe route. There are different ways of connecting pipes and are shown in Table 6.9. The only type of pipe connection that is visible on the P&ID is the flange.

However, flanges are not shown unless there is specific need for them.

6.14.1.5 End-of-Pipe Systems

End-of-pipe systems are applied to uncoupled pipes. Uncoupled pipes are the pipes that are not connected to other pipes and are not extended.

There are several available options for the end-of-pipe provision listed in Table 6.10. For pipe sizes less than 2", a screwed cap or plug can be used. The tendency is toward using a screwed cap in the newer plants. For pipe sizes more than 2", a blind flange or welded cap can be used. If there is a plan to extend the pipe in the future, a blind flange is generally used, but a welded cap is enough.

If a frequent connection to other hose is necessary, a quick connection is the best. Off-loading systems are using quick connections because the plan is frequent transferring of fluid to and from a transportation system.

Table 6.8 Rule of thumb for selecting three-way connections.


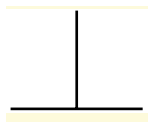
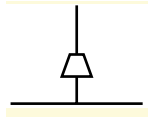

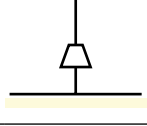
	If	Suitable three-way connection	P&ID representation
	Y is the same size of X	Tee	
	Y is the same is one size smaller than X	Reduced Tee	
	Y is smaller than X by two or three sizes	O-let	
	Y smaller than half of X	Tee and then reducer	

Table 6.9 Different methods of connecting pipes end to end.

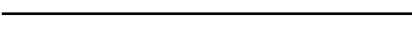
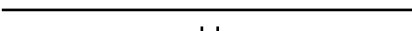
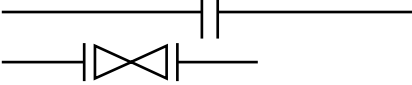
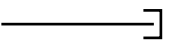
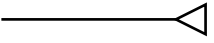

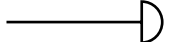
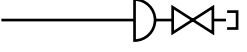
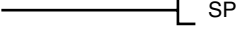
Component	Application	P&ID schematic
Socket welding, thread connecting	Pipe size less than 2"	
Butt welding	Pipe size more than 2"	
Flange		

Table 6.10 Different methods of ending pipes.

Component	Application	P&ID schematic
Screwed cap	Pipe sizes less than 2"	
Plug	Pipe sizes less than 2"	
Blind flange	Pipe sizes more than 2"	
Welded cap (with or without drain valve)		 
Quick connection	Tend to be used in small to medium bore size pipes	

6.14.2 Specialty Items

A plant is a combination of three hardware: equipment, instruments, and pipes. During the design of plant, the designers have an opportunity to design the equipment based on their process requirements. The instruments are selected based on the control requirement of the plant by instrument engineers. However, for pipe and pipe appurtenances, there is no such freedom. The process engineer must use whatever size of pipe or other piping items desired for the plant and put it on the P&ID.

The Piping group in each engineering company is responsible for purchasing piping and piping items, and they generally provide a list of all the acceptable standard types and sizes of piping and piping items that can be used in each project. This list is part of a document called a Piping Material Specification. Basically the Piping group tells process engineers that only items on the Piping Material Specification list must be used on a certain project. However, there are some cases that a process engineer or instrument engineer needs a piece of piping item, and it is not in piping spec. In such cases, the first option is to try to replace the desired item with another item in the piping spec. If that does not work, the Piping group may agree to this as an exception. These exceptions are considered SP items and definitely cannot be a long list. Therefore, it could be every non-equipment item that has not been included in the piping spec.

The Piping group does not care for SP items because they already know the complete specification of each item in the piping spec, and it is easy for them to buy the items from the market. However, for each SP item, the person who asked for a specific SP item (more than likely the process engineer) needs to prepare a data sheet for the item and submit it to the Piping group.

On the P&ID, a SP item is shown as a little box beside the item, and the acronym SP with a number, which is the tag number (Figure 6.86).

To decide if an element is an SP items, the piping spec of the project should be consulted. That also means that there is no universal rule that a non-slam check valve is a SP item. However, an experienced engineer may know that, generally speaking, where there is a check valve in the project's piping specs, it is a conventional check valve, and if non-slam check valves are required, they are

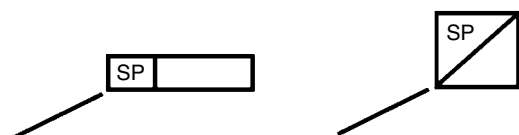


Figure 6.86 Specialty item tags.

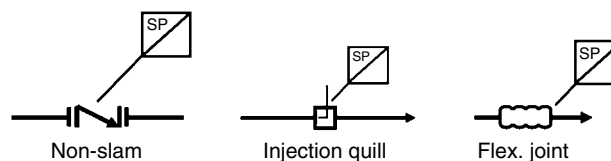


Figure 6.87 Specialty item examples.

most likely SP items. When a process engineer includes an element as an SP item, he/she needs to be aware that he/she does not have flexibility to choose the sizes needed as SP items are generally off-the-shelf items. Figure 6.87 shows few examples.

In this figure, a non-slam check valve, an injection quill for a chemical injection system, and a flexible joint are all SP items. In a P&ID, the designer may decide to put definition of the SP item beside the symbol. This is acceptable if the definition is one to three words in length. Otherwise it is better to not put the SP item definition on a P&ID and keep it only on the SP item data sheet.

Sometimes, there is a dispute between the Mechanical group and the Piping group if one specific element is SP item or actually equipment. This is a valid dispute because if the element is a piece of equipment, it should be tagged and bought by the Mechanical group, and if it is an SP item, it should be tagged and bought by the Piping group. As a rule of thumb, SP items should be small items on the pipe, meaning elements with less than 0.5m^3 in volume without any utility connection and without any complexity in its structure. If an element is small but it has a complicated structure, it is better to classify it as equipment rather than an SP item.

6.14.2.1 Flange-Insulating Gasket

Here, *insulating* refers to electrical insulation and not heat insulation. Galvanic corrosion is arguably the most common type of corrosion in process plants. Galvanic or electrochemical corrosion happens when two dissimilar metals are put in contact with each other. However, there are some cases that mating flanges are not similar from material point of view. These situations happen when there is a spec break in the pipe. The spec break is almost always on a flange. The flange could be for pipe, valve, or any other pipe appurtenance. One side of the flange is from one metallic material, and the other side is from another metallic material. In such cases, an electrically nonconductive gasket should be placed between two mating faces of the flange to prevent electrochemical corrosion. Flange-insulating gaskets can be tagged as SP items.

Table 6.11 Finding better items for replacing other items.

	Better choice	Worse choice
Deciding to put an element upstream or downstream of a reducer		
Deciding to put an element upstream or downstream of a centrifugal compressor		

6.15 Other Approach about Piping

In the ordinary approach, pipes are seen as a means of connecting equipment and containers to each other. However, the other approach sees pipe circuits in the background and equipment and other items placed on the pipe circuits. In this approach, whenever the P&ID developer decides to put an item on a pipe, it needs to be put on at the “best” place. The best place is decided based on technical and economic measures. Economic measures direct the P&ID developer to put the item wherever is cheaper. Inexpensive options could be a result of a smaller size item or items in more typical temperature and pressure ranges of operation.

This economical measure may show it is better to install a valve downstream of a reducer (smaller size) and not upstream, if the technical measures allows both of them (Table 6.11). This economic measure may show it is better to install a flow meter upstream of a centrifugal compressor (lower pressure) and not downstream, if the technical measures allows both of them (Table 6.11).

6.16 “Merging” Pipes

Pipes are not generally merged. Pipes within process plant are almost always for a dedicated service. The only important exception is by batch, by semi-batch, and by intermittent operations. In such operations, one single pipe route may be used for more than one service. For example, in sand filters, there could be a piece of pipe that during normal operations carries the dirty water, and during the back wash, the same piece of pipe carries backwash water.

For a pipeline, it is another story. In pipelines, it is common to see merged pipes, meaning a single pipeline may be used to carry different fluids in different time slots rather than having a dedicated pipeline for each

single fluid. The high cost of developing pipelines justifies using it as shared resource. The important thing when using a pipe or pipeline as a shared facility is the cleaning after each use. Without cleaning, there is a chance of contamination of transferring fluids in pipes or pipelines.

6.17 Wrapping-Up: Addressing Requirements of Pipe during the Life Span

In this section, the design is checked to make sure all the needs of pipes during each phases of plant life are functional. As it was discussed, these phases are: normal operation, non-normal operation, inspection or maintenance, and operation in the absence of one item.

- 1) Normal operation of pipe: The required considerations are already covered.
- 2) Nonroutine operation (reduced capacity operation, start-up operation, upset operation, planned shutdown, emergency shutdown): This phase of the plant needs two components to be handled – process and control and instrumentation. The process component is discussed here, but the control and instrumentation component will be discussed in Chapters 13–15. Generally speaking, pipes do not need to be considered much during the P&ID development to cover nonroutine operations. Pipes are robust items that can handle different conditions easily.
- 3) Inspection and maintenance: General consideration regarding inspection and maintenance of all items will be covered in Chapter 8. Here, however, we cover the specific requirements.
- 4) Pipe are robust items and do not need much attention during the operation.

Table 6.12 Bulk solid transfer methods.

Bulk solid transfer method	Explanation	Example
Pneumatic transferring in pipes	Transferring small size, nonsticky solids by air	Transferring gypsum powder
Hydraulic transferring in pipes	Transferring small size of solids by air	Transferring resin beads in ion exchange systems with <i>ex situ</i> regeneration
Hydraulic transferring in channels	Transferring medium to large size of solids by air	Transferring trees in rivers, Tomato transferring in tomato-processing plant
Screw conveyor	Transferring very small size, nonsticky solids, for flowable solids	Transferring starch
Belt conveyor	Transferring solids, not necessarily nonsticky	Transferring solid sulfur
Bucket elevator	Transferring large to very large size of solids by buckets	Transferring ores

- 5) Running plant in absence of pipe: There is generally no consideration for the times that a piece of pipe is out of operation because pipes are less prone to breaking. When a piece of pipe is out of operation, sometimes a temporary hose is placed to handle the same work.

6.18 Transferring Bulk Solid Materials

However, transferring solid material is more difficult than transferring liquids and gases. Because of that, there are plenty of methods to transfer solids, which is suitable for a limited range of applications.

Transferring solids is more common in plants whose raw materials are not liquids. The examples are mineral processing plants and food processing plants.

In the other plants, solid transferring may be used if the product is solid or there are materials in solid form.

The first attempt is to design the plant in a more vertical arrangement to transfer the solid by gravity. That is why it is not unusual to see a solid processing plant installed in stacked form or with inclined chutes. A common example is grain processing facilities. This is the best solution to minimize the use of or ease the use of solid-transferring equipment.

Table 6.12 is a nonexhaustive list of bulk solid material transfer methods. Because solid materials do not go through the equipment by themselves, plenty of these solid-transferring equipment are used as part of other equipment, for example, belt conveyor–dryer, screw conveyor–smasher, etc.

Reference

- 1 Stephen J. Emery. “Protecting against backflow in process lines.” Nov. 28, 1990 Chemical Engineering Magazine.

7

Manual Valves and Automatic Valves

Valves are a type of pipe appurtenances. Other pipe appurtenances, fittings, and specialty items were already discussed in Chapter 6.

7.1 Valve Naming

Valves are named based on their action on the service fluid (e.g. throttling valve, stopping (blocking) valve, or diverting valve) or based on their operating mechanism (e.g. motor-operated, solenoid, or manual valve).

Valves can also be named based on their plug type; they can be a gate, globe, or butterfly valve. Valves can also be named based on the valve's location in the piping (e.g. foot or root valve) or on its duty in the process (e.g. shutdown, blowdown, or flow control valve).

7.2 Valve Functions

Valves are piping appurtenances that actively affect flows. *Active* means the valve has a movable part. For example, a gate valve has a moving stem, and a check valve has a moving flap.

Any valve with moving mechanism that does not take order from the outside of the valve can be categorized as special valve. Check valves, air release valves, and excess flow valves are examples of special valves. Special valves are a large group of valves that have specific function. A few subgroups of special valves will be discussed at the end of this chapter. Special valves are diverse, which makes it difficult to classify them, but other valves can be easily categorized.

7.3 Valve Structure

Valves have two main parts: the operator and the plug (see Figure 7.1). The valve operator is the part that takes orders from an external source to act on the fluid flow.

The external orders could be an operator's hand or a nonhuman actuator. The valve plug is the part of the valve whose internal portion is in contact with the fluid. A valve plug can be built such that the valve is designed either as a throttling valve, an isolation valve, or a diverting valve.

The valve operator is connected to the valve plug by a piece of rod called the stem. The stem can move up and down or can turn depending on the type of the valve.

It is important not to get confused with *plug*: a plug is a part in all valves and a plug valve is a specific type of valve too.

7.4 Classification of Valves

There are at least three ways to classify valves: based on the action of the service fluid and their functions, based on the number of ports, and based on the number of seats.

The function of non-special valves in processing industries could be any of the following:

- 1) Adjusting flow valves
- 2) Stopping flow valves

Valves that adjust flow are valves that change the magnitude of flow in the pipes. These valves are called *throttling valves*. In these valves the valve plug is responsible for changing the magnitude of flow. In theory the valve plugs can take infinite position between fully open and fully closed, and hence valves can be categorized as either partially closed or partially open valves. The movement of the stem can change the plug position manually by an operator or via a remotely operated operator. These so-called valve operators will be discussed later in this chapter.

On the other hand valves that stop flow are valves whose plug can only remain in fully closed or fully open position. In both positions the valve allows a stream to flow. Therefore these valves are called *blocking valves*.

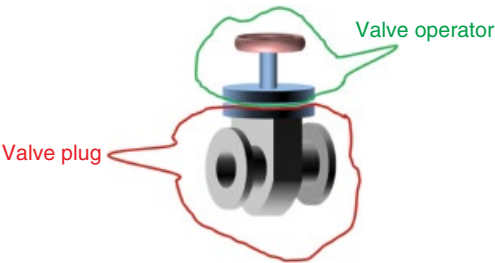


Figure 7.1 Structure of a typical valve.

A valve is either throttling or blocking based on the structure and shape of the valve’s plug. These plugs are discussed later in this chapter.

When placing a non-special valve in a system, the first question is if a throttling or blocking valve should be used. Since each calls for different types of valves, a wrong decision may cause internal valve leakage (passing-by) or premature failure of the valve. The design process engineer can decide whether the suitable duty of valve is throttling or blocking for a specific situation.

Furthermore valves can also be classified according to the number of ports: two-port or multi-port valves. The most common valves are the two-port valves, which have only two ports (i.e. one inlet and one outlet). However, multi-port valves have more than two ports, possibly three or four. They also have more than one inlet or outlet ports. Also they can be called by various names depending on the function of the valve.

Table 7.1 shows the four types of valves and their specific function. Based on the table a two-port stopping valve

Table 7.1 Valve action vs. the number of ports.

	Conventional (two-port)	Multi-port
Stopping flow	Blocking valve	Diverting valve
Adjusting flow	Adjusting valve	Adjusting-diverting valve

Table 7.2 Features of throttling vs. blocking valves.

	Throttling	Blocking
Application	Throttling	Stopping (blocking, on/off)
Stem travel: appropriate positions	0–100% wide (20–80%)	Zero OR 100% wide
Passing by?	Generally no tight shutoff	(Could be) tight shutoff
Example	A fully open valve may allow, for example, 45 m ³ h ^{−1} of flow; then a fully closed valve creates zero flow	A fully open valve may allow, for example, 45 m ³ h ^{−1} of flow; a partially closed valve may allow, for example, 30 m ³ h ^{−1} of flow
Interchangeability?	Can also be used in blocking applications	Cannot be used in throttling applications

is called an *isolation valve*. A two-port adjusting valve is called an *adjusting valve*. A multi-port stopping valve is called a *diverting valve*. A multi-port adjusting valve is called an *adjusting-diverting valve*. The two-port valves are by-default valves, and although the term “two-port” is not stated, it is generally assumed to be two-port valve.

Diverting valves are valves that divert the flow from one destination to another. Each diverting valve can be replaced with two or more blocking valves in a specific arrangement. Adjusting-diverting valves are valves that divert a portion of flow from one destination to another. Each adjusting-diverting valve can be replaced with two or more throttling valves in a specific arrangement.

Multi-port valves have no advantage over the two-port valves except saving space and money. They are not as robust as two-port valves. Through the use of multi-port valves, a huge cost saving is gained specially if using remotely operated valves (ROT). The remotely operated valves will be discussed later, but in a nutshell, they are actually valve actuators. When it is supposed to use remotely operated valves, merging few of them together and using multi-port valve generates a big saving as the saving in valve actuators are big.

7.4.1 Valve Plug: Throttling vs. Blocking Valves

Throttling valves can adjust the flow anywhere from 0% (fully closed valve and no flow) to 100% (fully open valve and full flow). However, the performance of the majority of these valves is best when they operate between 20 and 80% of flow.

In contrast, blocking valves allow for full flow or no flow at all. Although using a blocking valve in a throttling application is possible, it is detrimental to the valve internals in the long term. Throttling valves can also be used for isolation purposes, and although this does not damage the valve internals, they do not generally provide a tight shutoff (TSO). Blocking valves can be purchased as a TSO type. The concept is outlined in Table 7.2.

A valve could be throttling type or a blocking type depends on the plug structure. A valve can be named gate, globe, or butterfly type but each of them belongs to one of the groups of throttling or blocking valves.

Gate valves are the workhorses of the industry as blocking valves. They have a “gate” that goes up or down and provides a blocking function. Other blocking valves are ball valves and plug valves. Ball valves are inherently blocking valves. These have a ball with a cylindrical hole that rotates in a room. Plug valves can be considered the grandfather of ball valves. They are similar to ball valves, but they have a truncated cone inside instead of a ball.

Globe valves are the type of valve used in the industry for throttling applications. In globe valves a disk adjusts the flow that is perpendicular to it. Other types of throttling valves are V-port ball valves and characterized plug valves.

By changing the shape of the cylindrical hole to a V-port hole, a ball valve can be turned into a throttling type. Using this trick, an inherently blocking-type ball valve can be converted into a throttling valve. By cutting a “characterized” or “segmented” hole, a new throttling valve can be created. This can be seen in Figure 7.2.

The same trick, that is, cutting a characterized hole in the plug of a plug valve, can be used to convert a blocking plug valve to a throttling plug valve.

If the bore through the sphere is the same size as the nominal size of the valve, the valve is a full port ball valve. If the bore is one pipe size smaller than the nominal size, it is a standard port ball valve.

(3) Butterfly valves have characteristics of both throttling and blocking valves, and they can be used for both purposes; however, because they have high passing-by (internal leakage), they are not recommended for blocking purposes, especially in high-pressure streams or dangerous fluid service (flammable, toxic, etc.). A butterfly valve does work better as a control valve when the surface of its disk is not flat and is carefully contoured. Therefore, contoured butterfly valves are excellent as throttling valves.

Table 7.3 outlines a non-exhaustive list of valves in two applications in process plants: blocking and throttling.

On the P&ID each valve has a symbol that defined on the legend sheet of the P&ID set. However, the symbols



Figure 7.2 Two different plugs in ball valves that make it a blocking or throttling valve.

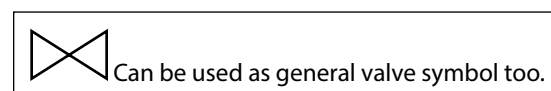
Table 7.3 Non-exhaustive list of valves.

	Blocking	Throttling
Workhorse	Gate valve (Conventional) ball valve Butterfly valve	Globe valve (Full port) ball Valve Butterfly valve V-port ball valve Characterized plug valve
Others	Plug valve	Pinch valve Needle valve Diaphragm valve

Table 7.4 Valve symbols on P&IDs.

Blocking	Symbol	Throttling	Symbol
Gate valve		Globe valve	
Ball valve		Butterfly valve	
Plug valve		V-port ball valve	
Butterfly valve		Characterized plug valve	
Pinch Valve		Needle Valve	
Diaphragm valve		Ball valve (full port)	
		Angle valve	

are different from each other in different companies. A typical list of valve symbols is shown in Table 7.4.



It is important to know that valve functionality, throttling or blocking, is not a stepwise characteristic; they are rather fuzzier features, meaning a gate valve is not purely a blocking valve and a globe valve is not purely throttling valve. This means that a gate valve has more blocking characteristics than throttling characteristics and a globe valve has more throttling than blocking characteristics. With that in mind, it could be said that conventional butterfly valves have fairly the same level of blocking *and* throttling characteristics. This concept is outlined in Figure 7.3.

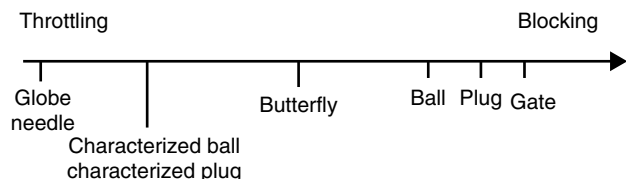


Figure 7.3 Valve spectrum regarding their functionality.

In majority of cases, a blocking valve can be considered a throttling valve only by limiting its range of operation. It is generally said that butterfly valves work well as valve in 30–70% of their range.

7.4.2 Valve Selection

There are plenty of resources to help designers find the appropriate type of valve. A quick-and-dirty outline to find the most suitable type of valve for a specific application follows.

The first step is deciding if the valve should be blocking or throttling. The process designer decides the type of valve (blocking or throttling) based on the expected functionality of the valve. Generally speaking, the majority of manual valves in process plants are blocking types. If there are manual throttling valves, they are mainly on non-primary pipes. The automatic valves (or remotely operated valves) could be either blocking or throttling depending on the expected function of the valve.

Figures 7.4 and 7.5 show the rules of thumb for deciding on the plug type of valves based on required size and the environment.

Generally speaking, a gate valve is used as default choice for a blocking valve, unless it is for very small sizes (e.g. 3") that a ball valve is used because miniature gate valves are expensive. Plug valves could be used in all the ranges, but they are not as common as gate or ball valves.

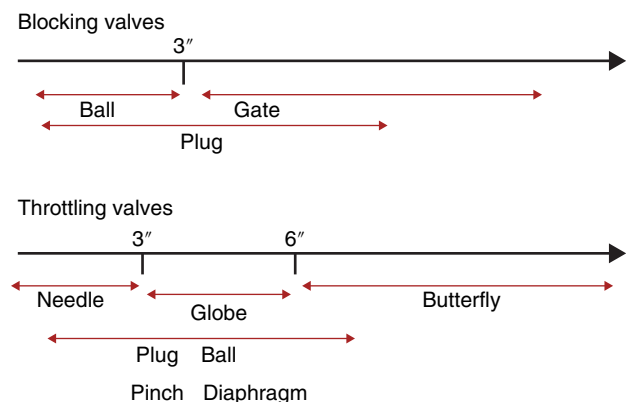


Figure 7.4 Rule of thumb for valve selection based on functionality and size.

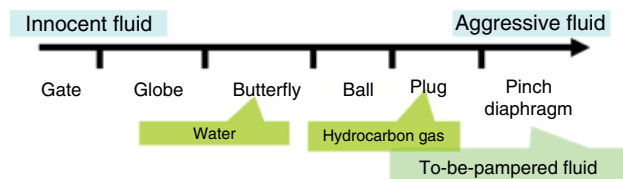


Figure 7.5 Rule of thumb for valve selection based on environment.

A globe valve is used for a throttling valve, unless it is for large sizes (e.g. 4" or 6") that a butterfly valve is used because large globe valves are expensive. In smaller sizes a specific type of globe valve, a needle valve, is used. When a high pressure drop is needed (e.g. more than 1000 KPa), a specific type of globe valve, an angle valve, can be used.

Characterized plugs, characterized balls, pinches, and diaphragm valves can be used in all ranges depending on service and availability.

Figure 7.5 shows the rule of thumb for deciding on the plug type of valves based on service fluid. There are a wide variety of available options when the service fluid is nonaggressive or "innocent." Butterfly valves are common in water services, and plug valves are commonly used for hydrocarbon gases. There are two cases in which a specific type of valve should be used: aggressive fluid (e.g. acids, slurries) and uncontaminated fluids (e.g. streams intended for human consumption).

The most paramount thing when dealing with aggressive or uncontaminated fluids is that no contact between valve internals (valve trim) and the service fluid is allowed. Pinch valves and diaphragm valves are the two types of valves suitable for aggressive and innocent fluids. In a pinch valve the service fluid is completely isolated from the valve internals by directing it through an elastic pipe inside the valve. The diaphragm valve is similar to a pinch valve (in this respect), but only partially isolates the service fluid from the valve internals.

Temperature and pressure are other factors to consider in making a decision on the plug type of valve. If a valve has polymeric elements (i.e. diaphragm, pinch, and sleeved plug valves), it has a temperature limitation on its application. Deciding which valve to use is dependent on the valve operators, which will be discussed later in this chapter.

7.4.3 Multi-port Valves

Multi-port valves could be either a blocking or throttling type. Multi-port blocking valves are better known as *diverting valves*. Throttling-type multi-port valves are also available, but they are less common than blocking-type multi-port valves.

Multi-port valves are not generally listed in a company's piping spec document. If this is the case, it should

be tagged as specialty item on the P&ID. Three- and four-port multi-port valves are the most common types, and they are generally available in sizes less than 6" at affordable prices. Multi-port valves tend to be used in

cleaner services. Symbols for three- and four-port valves are indicated in Table 7.5.

The blocking-type multi-port valves are generally plug, ball, or globe types. The plug route of multi-port valves is different, which allows them to provide different diverting functionality. The plug route is defined by the way how a plug redirects the flow. Three-port valves are uniquely designed in L port plug. The L port plug provides flow diversion in L shape (Table 7.6).

Four-port valves are made in different plug routes: straight port, L port, T port, and double-L port (Table 7.7).

Table 7.5 P&ID symbol for three-way and four-way valves.

Name	Symbol
Three-way valve	
Four-way valve	

Table 7.6 Different types of three-way valves.

Name	Internal	Position 1	Position 2
"L" port			

Table 7.7 Different types of four-way valves.

Name	Internal	Position 1	Position 2	Position 3	Position 4
"Dash" port					
"L" port					
"T" port					
"Double-L" port					

Generally the plug route of multi-port valves is not shown; however, if the multi-port valve is used in a critical application, it should be mentioned as a note beside the valve symbol or in the notes area of the P&ID. An example of using four-port diverting valve is shown in Figure 7.6.

It is shown in the figure that this arrangement reverses the flow of cooling water to the heat exchanger. The operator can reverse the flow every few weeks to reverse the flow of the tube sides to remove fouling from the tube internals. Probably using a multi-port valve in this application is the best because of the following:

- The service fluid (i.e. cooling water) is a non-dirty service to clog the valve.
- The reverse flow practice initiated every few weeks is not frequent and does not justify the use of several valves instead of one multi-port valve.

The multi-port valves are available for throttling purposes as well. The throttling-type multi-port valves are generally globe types. Multi-port throttling valves can be used for combining (mixing) or diverting purposes (Figure 7.7). One example of using multi-port throttling valves is shown in Figure 7.8. In this application the valve is used to adjust the flow of two streams that are combining. This valve can be replaced with two conventional valves that work in a parallel control system. The parallel control system will be discussed in Chapter 13. Some

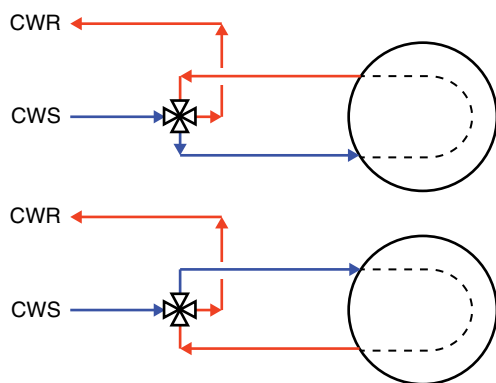


Figure 7.6 Application of four-way blocking valve on cooling water heat exchanger.



Figure 7.7 Throttling three-way valve.

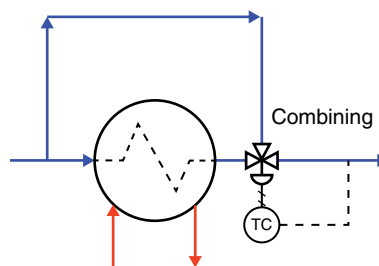


Figure 7.8 Application of three-way throttling valve for heat exchanger control.

designers may decide to use only one conventional control valve on the bypass stream. This is acceptable because between two streams, the bypass stream is the lower resistance route and more flow goes through it. Therefore, a single control valve on that stream satisfies the control goal.

7.4.4 Double-Seated Valves

In P&IDs double-seated valves can be shown as a specialty item and may use a special symbol.

In reality double-seated valves are distinguished from their respective conventional single-seated valves because of their bulkier body. In double-seated valves the stream is split into two streams and then goes through a dedicated seat for each stream. The main reason for using double-seated valves is to reduce the torque required to open or close the valve. A double-seated valve acting as a control valve needs a smaller pneumatic actuator in comparison with its respective control valve.

Double-seated valves, however, have an inherent problem, which is their passing-by. Because of the complexity in manufacturing double-seated valves, they almost always suffer from internal leaks or passing-by. Therefore double-seated valves rarely produce a TSO.

Double-seated valves can mainly be used in control valves on high-pressure streams when there is not enough room for a large pneumatic diaphragm (e.g. in debottleneck projects) as long as internal leaks are not a problem.

7.5 Valve Operators

There are two groups of valve operators: manual and automatic. Manual valve operators are the valves that can be field adjusted by an operator, whereas automatic operators that are installed on ROT are the valves that are controlled remotely from the control room. Automatic operators are also known as *actuators*.

The type of valve operator, either manual or automatic, totally changes the way it is handled in a process plant.

A manual valve is in the domain of the piping group, and during the design stage of a project, it is tagged (if it is supposed to be tagged) and procured by the piping group.

A remotely operated valve is listed, tagged, and tagged by the I&C group. Manual operators can be handwheel, gear operated, or chain operated. When a manual valve is chosen, a handwheel type is selected by default. The gear- and chain-operated valves are used in special cases.

When the valve flange size is large, it is difficult for the operator to open or close it. Therefore gear-operated valves are used in large valve sizes. Generally, a valve with a rotational stem can go from fully open to fully closed (or vice versa) after three times of nominal body-size rotation of the stem. It means that a 4" gate valve needs about 12 rounds of rotation to be in fully closed position from fully open position. When the valve flange size is equal or larger than 8", a gear-operated valve is used, whereas for valve flange sizes of 6" or smaller, the conventional handwheel valve can be used.

Chain-operated valves are used if a valve is not accessible to an operator. When a valve is elevated more than 2 m, it is said to be inaccessible. The solution is to install a platform or catwalk or use a chain-operated valve. Installing a platform or catwalk (not visible on a P&ID) is an expensive solution and is justifiable only if there are many elevated items (including valves and other pieces of equipment) that can be accessed by the catwalk. In other cases, only a chain-operated valve can solve the problem of an elevated valve.

It is important to know whether a valve is gear operated or chain operated; it is most likely be tagged as specialty item on the P&ID because the majority of piping specs list only handwheel manual valves.

Deciding between manual (cheaper) and automatic (more expensive) valves depends on the type of valve plug (blocking or throttling) and also on the situation. For a throttling valve to adjust the flow only once in the life of the plant, there is not really a need to use a throttling valve, so only a restriction orifice (RO) satisfies the requirement. An RO is less expensive than a valve and is basically a donut-shaped plate installed in a pipe to restrict the flow. The RO acts as a throttling valve maintained on a specific valve opening. An RO is tagged as a specialty item on a P&ID. If a throttling valve needs to be operated a few times a month, a manual throttling valve will probably work best. However, a throttling valve that needs to be changed to different openings several times in a day or even several times an hour cannot be manual type, and an automatic throttling valve or control valve is used instead. This is shown in Figure 7.9.

There are also cases in which a process plant is an arid and harsh area and expecting an operator to go the field and change the valve position even a few times a month could be unrealistic, and a throttling valve should be used.

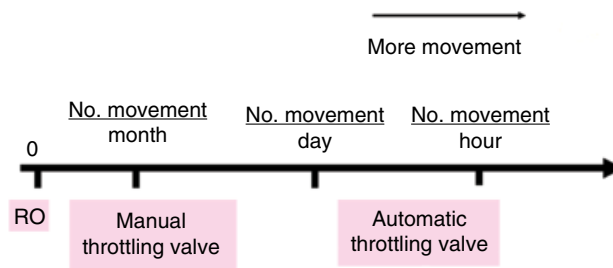


Figure 7.9 Rule of thumb for selection of operators for throttling valves.

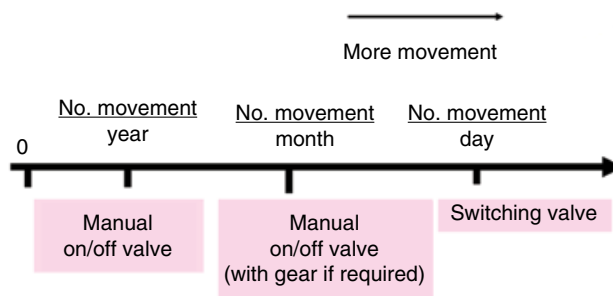


Figure 7.10 Rule of thumb for selection of operators for blocking valves.

Also when dealing with toxic material, the designer may decide to use an automatic operator for the majority of valves to minimize operator contact with plant facilities.

For a blocking valve, a valve with a one-time change in the life of a plant basically means nothing in the pipe (if the valve supposed to be open always) or a blinded pipe (if the valve supposed to be closed always). Therefore, in a case with no change, a blocking valve should not be used. If a blocking valve needs to be opened or closed few times in a month or a year, a manual operator is enough. However, if a blocking valve needs to be change in the plug position several times in a day, an automatic operator may be used. This is shown in Figure 7.10.

Generally speaking, there are more manual blocking valves in process plants than manual throttling valves. There are fewer manual throttling valves because operators are needed to change the flow and relying on operators is not in the best design of the process. There are more manual blocking valves not because they can be in process roles or isolation roles (isolation will be discussed in depth in Chapter 8). Manual valves are more common in small plants than medium or large process plants.

7.6 Different Types of Actuators

To build an automatic throttling valve (or control valve), a throttling manual valve should be attached to a modulating actuator. And to build an automatic isolation

Throttling valve + Modulating actuator = Control valve

Blocking valve + On/off actuator = Switching valve

Figure 7.11 Control valve and switching valve.

valve, an isolation valve should be matched with an on–off actuator.

Automatic throttling valves are called *control valves*, and automatic blocking valves are called *switching valves* (Figure 7.11). There are different types of automatic actuators available on the market. Table 7.8 summarizes their features. It can be seen from Table 7.8 that the majority of actuators are of the modulating type. The only actuator that is mainly an on–off type is a solenoid actuator. However, modulating solenoid actuators are available but are not common. The most common actuators in process plants are diaphragm actuators.

7.7 Basis of Operation for Automatic Valves

An automatic valve can be triggered by time, event, or human action. Time can be the basis for the operation of an automatic valve, but this is not true for throttling

valves. An automatic isolation valve can be controlled with time, and it means that this valve can be scheduled opened or closed at specific times and for a specific duration.

This type of application is common for switching valves in semi-continuous or intermittent operations. For example, in a filtration system, the filter is in normal service (performs filtration) for 23 hours, and then it needs to be pulled out of operation for backwashing for half an hour. In this example, multiple switching valves around the filter vessel are scheduled based on time to handle the operation.

Event-based automatic valves are automatic valves that take orders from a sensor signal. This type is common for both throttling and isolation valves. All control valves in control loops are event based. The majority of switching valves in emergency shutdown system are event based; this means that if the pressure in a vessel goes beyond a specific number, a switching valve will be opened to release the pressure.

The last basis of operation is human action, meaning the operator is in the control room, not in the field; this means the valve takes orders from a remote operator. For throttling valves, a remotely operated valve is referred to as a hand control valve. For blocking valves, a remotely operated isolation valve is called a shutoff valve.

Table 7.8 Different types of actuators.


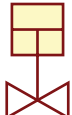
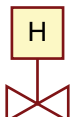
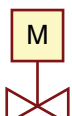
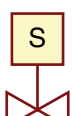
Type	Driver	Acronym	Type	Symbol	Application
Diaphragm actuator	Instrument air	AOV	On/off or modulating		The default choice
Piston actuator	Instrument air		On/off or modulating		Where the diaphragm type cannot be used (e.g. the required diaphragm area is too large)
	Hydraulic oil	HOV	On/off or modulating		Where instrument air (IA) is not available or pipe pressure is high
Motor-operated actuator	Electric motor and AC electricity	MOV	On/off or modulating		Where instrument air (IA) is not available. For large bore pipe size (e.g. >12")
Solenoid actuator	Solenoid and DC electricity	SOV	Almost always on/off		Mainly for sizes less than 2", generally on instrument air tubes (not common for process pipes)

Table 7.9 Two main types of valves and their tagging.

	Blocking		Throttling	
	Name of valve	Tag	Name of valve	Tag
Time	Sequence valve	KV	–	–
Event	Switching valve	XV or UV	Control valve	FV (flow control valve)
Human operator	Shutoff valve	XV	Hand control valve	HCV (hand control valve)

7.8 Tagging Automatic Valves

Generally speaking, all ROT should be tagged on P&IDs. Table 7.9 summarizes the methodology for tagging automatic blocking valves and throttling valves.

7.9 Tagging Manual Valves

Tagging manual valves is not as common as tagging automatic valves. Some companies do not tag their manual valves because they consider manual valves to be less important than automatic valves. However, in process plants that produce critical products, such as nuclear power plants, explosive material plants, and herbicide production plants, it is common to see tagged manual valves.

The manual valve tag anatomy is generated on a per-company basis, but it generally has an alphabetical acronym and a sequential number.





7.10 Valve Positions

When talking about valve positions, there are two different positions: regular and fail. Regular position is the position during normal operations of plant, and the fail position is the position when losing the driver of “operator.”

The regular position of a valve is an attribute of blocking valves, either manual or automatic. Regular position refers to whether the valve is fully open during normal operation of the plant. To show this position, an acronym is placed under the valve on the P&ID. For example, NO means normally open. When a valve is NO, this means that during normal operation of the plant, the valve should be fully open, closing only during plant maintenance, shutdown, and so on.

The fail position of a valve is an attribute of automatic valves (either throttling or blocking). It refers to the position of the valve when the “driver” of the valve is lost (fail condition). For example, a control valve (automatic throttling valve) could be called fail open (FO); this

Table 7.10 Regular and failure position of valves.

	Throttling	Blocking
Manual		 NC
Automatic	 FC	 FC NC

means that if the instrument air to this valve is lost, then the valve will go to its open position and stop in that position. Table 7.10 shows the assigning of the regular position and failure position to the respective valves as a clarification to the preceding discussion.

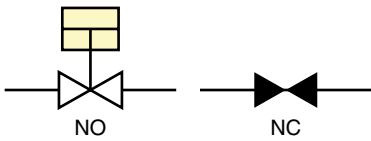
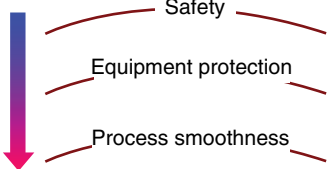
In the next two sections, regular positions and failure positions will be detailed more fully.

7.10.1 Regular Position of Blocking Valves and Decision Methodology

Regular positions of blocking valves are classified in three pairs: normally open/normally closed, locked open/locked closed, and car seal open/car seal closed. Each could be used in a condition, but all pairs represent a valve that has a regular position of being open or closed.

Deciding on the position of block valves is done by evaluating three items: safety, equipment protection, and process smoothness (Table 7.11). If a valve needs to be opened or closed just to support the process, it can be specified as normally open (NO) or normally closed (NC). If the decision for the valve’s regular position is based on equipment protection, then the valve would be locked open (LO) or locked closed (LC). If a blocking valve needs to be closed to protect a plant from a hazard, it is specified as car seal closed (CSC). If it needs to be open for safety reasons, then this valve is specified as car

Table 7.11 Regular position of blocking valves.

Only for:	Blocking valves; manual or automatic
Concept	Position in routine operation of plant
Examples	
Acronyms	<div><div>CSO, CSC</div><div>LO, LC</div><div>NO, NC</div></div> <div></div>

seal open (CSO). As a note, not all companies use these three elements.

To show the regular position of blocking valves, the corresponding acronym is stated under the valve symbol. When a valve is NO, this means that during normal operation of the plant, the valve should be fully open, closing only during plant maintenance, shutdown, and so on.

NO or NC is used only if there could be confusion around whether the valve should be open or closed during normal operation of the plant. All blocking valves are generally considered NO valves, unless NO is stated under the valve symbol. Instead of putting the acronym NC or NO under the valve symbols on the P&IDs, some companies use a filled symbol of valve to show it is a NC valve (rather than an outlined symbol). A good example of using NC valves for all vent and drain valves is when the vent and drain valves are closed during the normal operation of the plant and are used only for start-up.

Examples of using LO/LC and CSO/CSC will be shown in future chapters.

7.10.2 Failure Position of Automatic Valves and Decision Methodology

The fail position for an automatic valve could be fail open (FO), fail closed (FC), fail last (FL), and fail indeterminate (FI) (Figure 7.12).

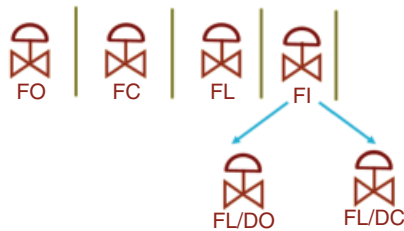


Figure 7.12 Failure position of automatic valves.

The acronym FL can be a source of confusion because some interpret it as fail last and others as fail locked. Fail last means that after losing power, the valve will stay wherever it was. If this is the concept, it is better to name it note it as FI because in reality, a valve that stays in its last position will drift to the open or closed position because of fluid pressure. If the meaning of FL is fail last, then drift open or drift closed should also be specified: FL/DO or FL/DC; otherwise, it will be interpreted as fail locked. Control practitioners should be aware of this and clarify the concept of FL to avoid confusion during the procurement or design of a plant.

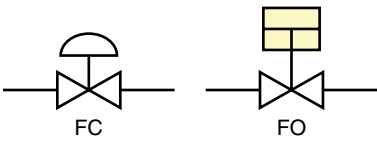
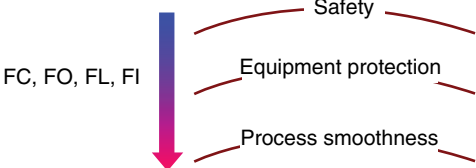
The same three elements used in deciding the position of blocking valves are used to decide the failure position of an automatic valve: safety, equipment protection, and process smoothness. Based on these elements, it will be decided if an automatic valve should be open (FO), closed (FC), locked in the last position (FL), or stops but not locked in the last position (or “last,” FL). For automatic valves, unlike block valves, the selected acronym (FO, FC, and FL) does not show if the governing decision was based on safety, equipment protection, or process smoothness.

It means if an automatic valve should be open, FO is placed below its symbol, no matter why the decision was made to keep it open.

To decide the failure position of a valve for safety reasons, the need to minimize the hazard in the plant is the most important. All valves located on high-energy streams should generally be FC. Examples of these streams are a fuel gas stream going to the burner, hot glycol, and steam. All valves located on low-energy streams should generally be FO. Examples of these streams are cooling water and nitrogen gas. The intent when determining the failure position based on safety is to decrease the energy of the plant by lowering temperature, lowering pressure, and so on.

Sometimes safety does not dictate the failure position of the valve, and in such cases, the designer may

Table 7.12 Failure position of automatic valves.

Only for:	Automatic valves; blocking or throttling
Concept	Position when losing “operator”
Examples	
Acronyms	

decide to arbitrarily pick a failure position, such as FC or FO (which are cheaper than FL), and assign it to the automatic valve. However, some clients ask the practitioner to use a second decision tool if safety does not dictate a failure position. In such cases, the second decision tool (equipment protection) can be used and then followed by process smoothness if needed. This means that if safety does not dictate the failure position of the valve, the designer should check what the best failure position of the valve would be to best protect the equipment.

A good example of deciding where to locate failure positions based on equipment protection is to make a control valve on a minimum flow line on a set of centrifugal pumps, all of which are FO. In this case, the decision is

made not based on safety, but to protect the pump. Whether this valve is FO or FC does not impact the safety of the plant; however, if it is FC, it will impact the integrity of the equipment or pump. This means that if the valve is FC and instrument air is lost, then the minimum flow line of the pump will be blocked and the pump will not be protected against low-flow cases and may be damaged.

Process smoothness is the last decision element if the first two (safety and equipment protection) cannot determine the failure position of the valve. Process smoothness basically recommends the failure position that is the best for the process and that minimizes the fluctuation magnitude and extent when a failure happens.

The concept of valve position is summarized in Table 7.12. When there is no constraint exerted by safety, equipment protection, or process smoothness, the failure position is decided based on economic factors; if FO and FC valves are the least expensive automatic valves, then one of them is selected. And sometimes the failure position is not mentioned in the P&ID at all.



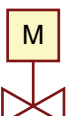
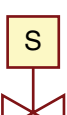
7.10.3 More Concepts about Failure Position of Automatic Valves

• What Is the *Failure* in Failure Position Discussion?

As was mentioned previously, *failure* means losing the driver. *Driver* is combination of all energy streams that direct the valve actuator (and consequently the valve stem) to a specific position. So, in this context, *failure* does not refer to mechanical failures or any type of jamming.

Table 7.13 shows the different actuators and their drivers.

Table 7.13 Failure reasons of different actuators.

Type	Symbol	Actuator driver(s)	Failure cause(s)
Diaphragm actuator		<ul style="list-style-type: none"> • Instrument air • DC electricity 	<ul style="list-style-type: none"> • Losing instrument air • Losing DC electricity
Piston actuator		<ul style="list-style-type: none"> • Instrument air • DC electricity 	<ul style="list-style-type: none"> • Losing instrument air • Losing DC electricity
Motor-operated actuator		<ul style="list-style-type: none"> • AC electricity 	<ul style="list-style-type: none"> • Losing AC electricity
Solenoid actuator		<ul style="list-style-type: none"> • DC electricity 	<ul style="list-style-type: none"> • Losing DC electricity

In some types of actuators (i.e. diaphragm and piston actuators), there are two energy streams as the actuator driver: instrument air and DC electricity. In these actuators, instrument air moves the main actuator element to push the valve’s stem. But the instrument air is initiated to the actuator through one solenoid valve or an arrangement of solenoid valves, which are functioning via DC electricity. If either of these gets lost, the actuator fails to function.

Losing instrument air is known as a power loss and losing DC electricity is known as a signal loss [1].

To what type of failure does the failure position of an automatic valve refer? Losing instrument air or losing DC electricity?

If failure position of an automatic valve is mentioned without any specifics, it is generally because of power loss. However, to prevent any confusion, it is better to

clearly mention if the failure position is for power loss or signal loss.

The letter P at the beginning of acronym for failure position shows it is for power loss, and an S represents signal loss (Table 7.14).

A process engineer generally wants to have an automatic valve with the same failure position for power loss and signal loss. Complications can arise if a design process engineer asks for a failure positions in signal loss differently than in power loss. The automatic valves generally report the failure position in cases of power loss unless another option is needed.

• **Actuator Driving System**

Showing the driving system of valve actuators on the P&IDs are different for each company. Some decide to show all the details of the driving mechanism, and others show only a brief schematic of the system and refer the reader to other documents for details of the system. Table 7.15 shows different ways of displaying automatic valves on P&IDs in regard to their driver system. Chapter 13 discusses actuator driving systems in more detail.

The symbol of a multi-port valve in the detail of drivers does *not* refer to process multi-port valves. These valves are common in the hydraulics industry

Table 7.14 Failure case acronyms regarding different types of driver loss.

Driver loss type	Name of drive loss	Representing acronyms
Instrument air	Power loss	PFC, PFO, PFL, PFI
DC electricity	Signal loss	SFC, SFO, SFL, SFI

Table 7.15 Automatic valves on P&IDs.

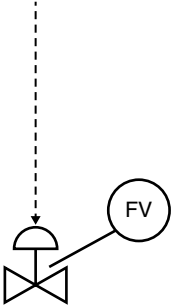
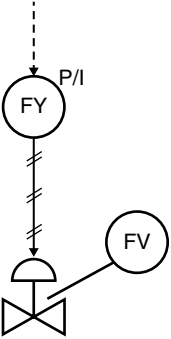
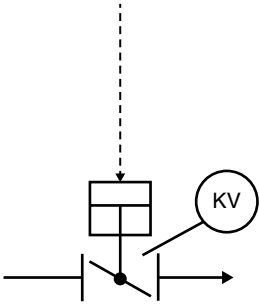
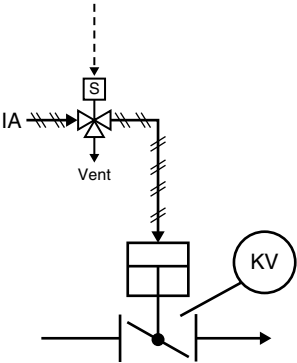
	Simplistic presentation	Detailed presentation
Control valve		
Switching valve		

Table 7.16 Explanation of two poppet valves.

Poppet valve name	Symbol (on P&ID)	Poppet valve structure
3/2-way valve		
4/3-way valve		

and are known as poppet valves. In the hydraulics industry, they have specific symbols, but on the P&IDs it is common to use the same symbols for (process) multi-port valves for the poppet valves, which could be confusing. Table 7.16 explains two poppet valves. These poppet valves are energized by solenoids, so they are a kind of solenoid valves.

The driving system of switching valves can be complicated depending on the required failure position of the valve. Manufacturing companies working with different arrangements of solenoid valves on the instrument air route and different types of actuators make automatic valves FO, FC, fail locked, or fail last.

Diaphragm and piston actuators, in this regard, are single-acting top spring, single-acting below spring, and double-acting diaphragm/piston. Single-acting actuators are the ones whose diaphragm or piston has one instrument air connection and comes with spring. The double-acting actuators have two instrument connections and may not have spring.

Here we do not go any further and just show few examples in Figure 7.13.

7.11 Valve Arrangement

Valves can be installed in single arrangement, in parallel, or in series.

A single valve can be installed in a piping system with its flanges attached to nuts and bolts or by welding or

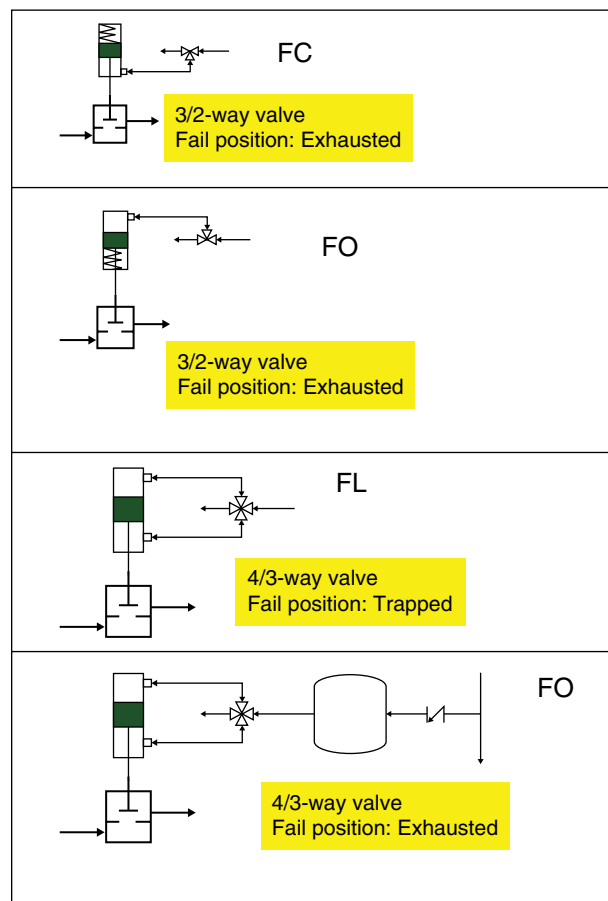


**Figure 7.13** Few examples showing switching valves with detailed driver.

Table 7.17 P&ID symbol for connecting valves.

Valve size	Connection type	P&ID sketch
Nominal size <2"	Weld	
Nominal size >2"	Flange	

screwing its ports by fitting it between the flanges of the two sides of pipes (Table 7.17).

Generally valves are installed between piping sides without any other fittings. However, one important exception is when installing control valves. Sometimes, the selected control valve has smaller body size that can be fitted with the pipe size. In such cases, a reducer on the inlet of a control valve and one enlarger on the outlet of the control valve may be needed.

7.11.1 Valves in Series

A manual valve can be used in series, one blocking type and one throttling type. If a stream needs to be adjusted manually, and sometimes the stream should be totally stopped and tight shutoff is important, it is a good idea to use a manual blocking valve and then manual throttling valve in series (Figure 7.14). This arrangement can be used in services like toxic fluids or high-pressure streams.

Two (or more) manual throttling valves or two (or more) manual blocking valves are rare, which sometimes is considered a bad practice in P&ID development. However, sometimes having two or more manual blocking valves in series happens. Each piece of equipment needs isolation valves around it for ease of maintenance. However, when there are two pieces of equipment, one upstream and one downstream, and they are close to each other, two of their isolation valves sit close to each other in a series position. In such cases, one isolation valve can be eliminated (Figure 7.15).

Valve arrangement in series can be used for control valves or regulators. These are for the cases that a large pressure drop is needed in a stream. A large dropping pressure may cause vibration, noise, and erosion in the valve [2].

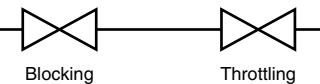


Figure 7.14 Manual valves in series: blocking and throttling.

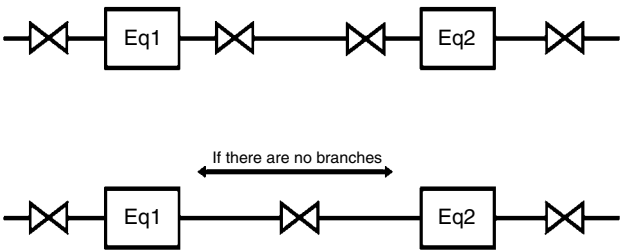


Figure 7.15 Two manual blocking valves in series and saving opportunity.

A rule of thumb helps to decide when two regulators in series may be needed:

- Where a pressure drop more than 100psig is needed (or maximum 150 psig).
- Where pressure should be dropped to a value less than 1/10th of upstream pressure.
- Where the pressure on downstream should be accurately regulated (e.g. less than few psig).

7.11.2 Valves in Parallel

The parallel arrangement of two manual valves may be used. There are some cases that a manual blocking valve needs to be placed on a stream that has high pressure. In such cases, placing one single blocking valve, for example, a gate valve, makes life hard for the operator who will have to open a manual valve from a fully closed position under high pressure (e.g. more than 3000 KPa). To solve this problem, another smaller-sized manual blocking valve is installed in parallel with the main valve. When the operator wants to open the main valve, the small bypass valve is opened at the beginning to equalize the pressure in both sides of the main valve, and then the main valve can be easily opened (Figure 7.16).

Parallel manual valves could be used for other reasons, such as providing a minimum flow in the pipe even when the main valve is closed or for start-up. As was discussed in Chapter 5, the general method of starting up a piece of equipment involves gradually opening the valve of the inlet stream. If the operator does not want to open the valve suddenly, a parallel and smaller manual valve can be added to the main valve and the operator can open it

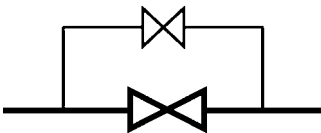


Figure 7.16 Two manual blocking valves in parallel.

(even quickly). In some designs, this small parallel valve is called a *warming valve*.

Control valves can be used in parallel. There are at least two cases in which a control valve may be used in parallel. One, as discussed previously, is as a spare control valve in parallel to an operating control valve, and the other one is using two control valves in a split-range or parallel control (to be discussed in Chapter 14).

Switching valves in parallel is rare, especially if the switching valve is responsible for shutting down a stream for safety purposes.

7.12 Control Valves and RO Combinations

A control valve can be combined with RO in two basic forms: in series and parallel. When an RO is placed in series with a control valve, it limits the maximum flow through the control valve when it is wide open. The same goal can be achieved by buying and placing a control valve with mechanical stop (Figure 7.17).

A control valve can also be limited in the wide open situation by implementing electrical or software stops through the control system. An RO in parallel to a control valve warrants a specific flow going downstream irrelevant of the control valve stem position, even when it is fully closed. The same goal can be achieved by placing a control valve with mechanical stop (Figure 7.18). A control valve can also be limited in the fully closed situation by implementing electrical or software stops through the control system.

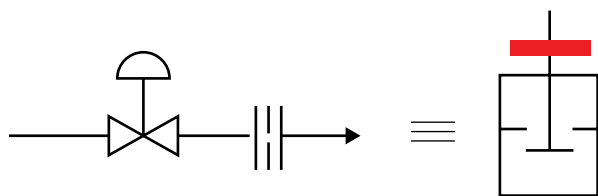


Figure 7.17 Similarity between “control valve in series with RO” and “control valve with mechanical stop.”

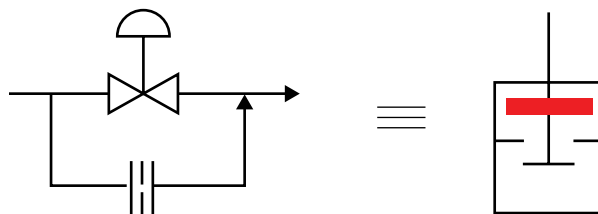


Figure 7.18 Similarity between “control valve in parallel with RO” and “control valve with mechanical stop.”

7.13 Operating in the Absence of Valves

Operating in the absence of valves, manual or automatic, will be discussed. It is rare to implement any provision for the times manual valves are out of operation. Generally, manual valves are robust and will fail not frequently. However, it is a different story for automatic valves. This concept is discussed in more detail, one section for control valves and other section for switching valves.

7.13.1 Operating in the Absence of Control Valves

The mean time between failures for control valves are shorter than a typical scheduled plant turnaround. Because of that, some provisions should be implemented in the design of control valve arrangement to make sure that there is always a controlling element available in the system. There are at least three different arrangements to satisfy this requirement.

The least reliable system is a single control valve on the pipe. In this case, only two isolation valves on each side of the control valve are installed to make sure that the control valve can be removed easily after closing the isolation valves. However, this arrangement obviously impacts the rest of the process plant when the control valve is not in place and not working. This arrangement is acceptable only in noncritical services like batch operations, for example, on a pipe that is in intermittent service of working 4 hours in 24 hours. In case the control valve failed, it can be removed from the system and get repaired and replaced within the 20 hours of nonoperation. In this case, such arrangement does not limit the reliability of the system (Figure 7.19a).

On the other side of the spectrum is an arrangement that provides the highest level of reliability. In this arrangement, two control valves are installed in parallel: one of them works as an operating control valve and the other one as a spare control valve. These control valves work on a sparing philosophy of $2 \times 200\%$. This is the best arrangement when dealing with critical services and when a high reliability of control valves is needed. However, it is expensive because one control valve with a complete control loop is kept as a spare. Note this arrangement is rare in industry (Figure 7.19b).

The third option is the one that provides a good balance of reliability and cost. This option is putting another valve in parallel to the control valve, but this valve is only a manual throttling valve. In this case, a spare is used for a control valve, but that spare is not a full control valve and is instead a simple inexpensive

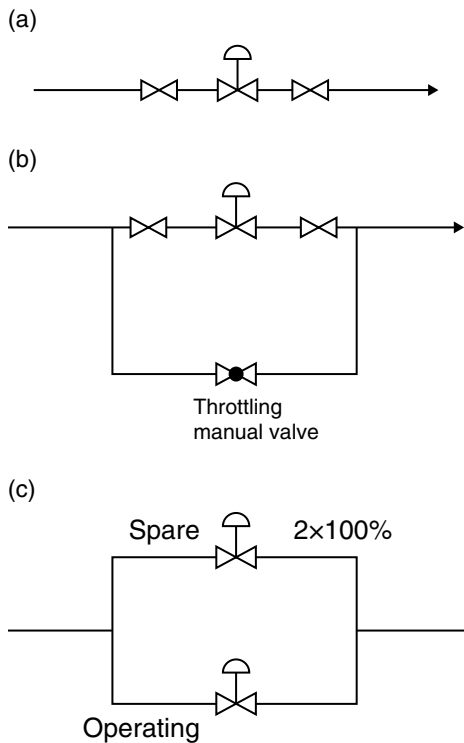


Figure 7.19 (a–c) Different arrangements for control valves to provide reliability.

manual throttling valve. This arrangement is common for control valves and is called control valve station (Figure 7.19c).

The basic arrangement of a control station is a control valve with two isolation valves on each side of control valve, and the bypass pipe is outside of the isolation valves with a throttling valve on it (Figure 7.20).

Whenever there is a problem in control valve wherein the control valve should be pulled out from the operation and inspected or be sent to the workshop for maintenance,

the isolation valves around the control valve are closed, and an operator stands up beside the bypass throttling valve and puts his hand on it while watching the sensor reading, the sensor that is used to send order to the under-the-maintenance control valve and adjust the opening percentage of the manual valve to mitigate the system. Here basically we used another control loop in the absence of the main control loop. However, this fake control loop is handled by an operator. Definitely an operator cannot work as a “control loop” for long period of time because it is very boring and tiring job. However, it is acceptable to ask an operator to take care of the control valve duty for short period of time when the main control valve is under maintenance.

It is obvious that the manual bypass valve is fully closed in majority of time during the life of a process plant. We know that throttling valves are not very reliable as a tight shutoff device. Therefore, in critical services, another valve could be placed upstream of the bypass valve to work as a dependable tight shutoff blocking valve. The example of such critical services could be toxic services, aggressive chemical services, or high pressure steams.

It will be discussed in Chapter 8 that venting and draining of the control valve station is important before performing any inspection or maintenance on it. Therefore, the operator should fully drain and vent the system. The important question is how to drain and/or vent this part of pipe to make sure there is no chance of liquid splashing or gas pushing during inspection and maintenance.

In plenty of cases the control valve flange sizes are smaller than the pipe sizes. Therefore reducer and enlarger are needed in the size of the control valve. If the pipe size is large, it is a good idea to use eccentric reducer and enlarger instead of less expensive concentric reducer/enlarger. This trick provides full drainage for the system. Full drainage of the system is important when the service is liquid or contained liquid or an aggressive fluid and

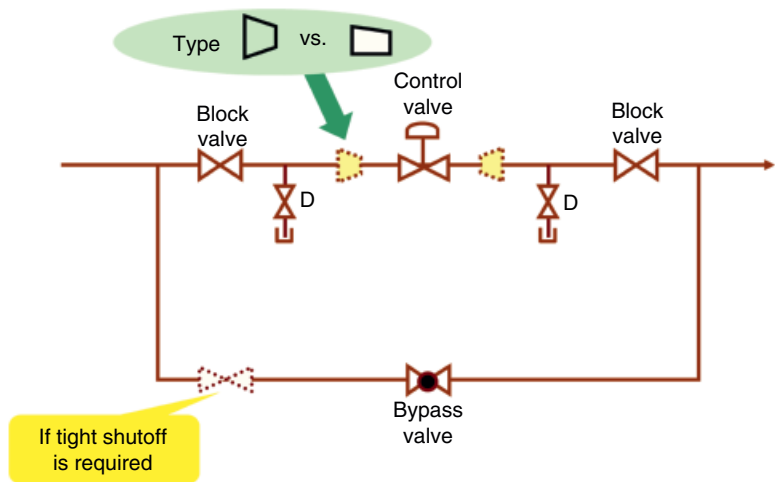


Figure 7.20 Details of control valve station.

when the pipe diameter size is large enough to contain a decent amount of liquid if concentric reducer and enlarger are to be used. Therefore, for cases that deal with aggressive liquid and large pipe size (say, more than 3 in.), it is a wise decision to use eccentric reducer and enlarger with flat on bottom (FOB).

The other provision for draining/venting the piece of pipe between two isolation valves is putting drain valve on the bottom of the pipe and vent valve on the top of the pipe. However, if the pipe size is small enough (say, less than 3 in.), possibly just one valve from the bottom of the pipe could be enough to be used as drain valve and vent valve.

Now the question is whether the drain valves are needed in both sides of control valve or just one side of that. There is a heated debate among professionals regarding this question. If there is an available guideline in your company you need to follow, otherwise you need to make the judgment whether to put two drain valves in one side of the control valve or just one drain valve downstream or upstream of control valve. There are at least three different answers for this question. A very conservative approach says that we need to put the drain valve in each side of the control valve to make sure that each side of the control valve can be drained independently. This approach is the best for critical cases like when dealing with toxic, hazardous, or high-pressure stream. This is also a good decision when a control valve is FC. In such cases where the control valve is FC, there are trapped liquids in each side of the control valve, one trapped liquid upstream of the control valve and the other trapped liquid downstream of the control valve. Therefore, two drain valves help the operator to drain each trapped liquid easily from each side of the control valve.

The other approach says that one drain valve for this piece of pipe is enough. Professionals who are in favor of this solution are faced with the following question: "what if the control valve failed or jammed in closed position?" They answer that the control valve could be opened by a jackhammer and again there will not be any two separate trapped liquid. Therefore one drain valve is enough. If we chose this approach, there are again two available options: if the single drain valve should be upstream or downstream of the control valve.

Some people prefer a single drain valve upstream of the control valve. They believe that a drain valve upstream of the control valve helps us to drain the higher pressure side of the control valve more safely and if it failed to open up a jammed closed control valve, the other side of the control valve has lower pressure and possibly does not need to be drained through the drain valve. The downstream of the control valve can be naturally drained after disassembling the control

valve and removing it from the piping arrangement. Putting drain valve upstream of the control valve has some other advantages for the operators. They can use this valve for start-up and for purposes of chemical cleaning. The other option is putting the single drain valve downstream of the control valve. This option also has some supporters.

To summarize the discussion about the need for drain valves on the control station, some people believe that we need to consider the failed position of the control valve. However, some other people do not take this into consideration on putting drain valve or drain valves around the control valve.

The last thing that should be decided is the type of bypass manual throttling valve. The workhorse of the industry for throttling valve is the globe valve. Therefore, wherever we want to put a bypass throttling valve for a control valve, a globe valve is selected. However, globe valves are not available (or are very expensive) in larger sizes, probably not larger than 4 or 6 in. Where the control valve size is larger than 4" or 6", there are some options available.

One available option is using butterfly valves. Butterfly valves are good throttling valves and are available and affordable for large sizes, for example, more than 4 or 6 in. However, not all companies and professionals are in favor of using butterfly valves. The conventional butterfly valves have some inherent drawbacks. They may have internal passing-by that makes them unsuitable valves for high-pressure systems and where the service fluid is an aggressive fluid.

If the required size of throttling valve is more than 4 or 6 in. and butterfly valve is not an acceptable option, there are some exotic designs available.

One completely acceptable option is providing the required valve capacity of the control valve through several small (less than 4 or 6 in.) globe valves. In this solution the bypass of the control valve could be an arrangement of two (or more) 4" manual globe valves. This arrangement – shown in Figure 7.21 – is fully functional but is very expensive.

If it is known that the movement of the control valve stem is only in a short and limited span, a manual globe valve can be replaced with a gate valve and a smaller globe valve in parallel (Figure 7.21 top schematic).

However, it has been seen in cases that a company decided to put a gate valve as the bypass valve for a control valve. As gate valve is NOT a throttling valve, it is not a good choice in this situation. However, a gate valve in parallel to a manual globe valve can be considered if it is discovered that the control valve mainly works on its extreme sides of its range.

At the end it should be mentioned that some companies and professionals believe it is not a good idea to provide

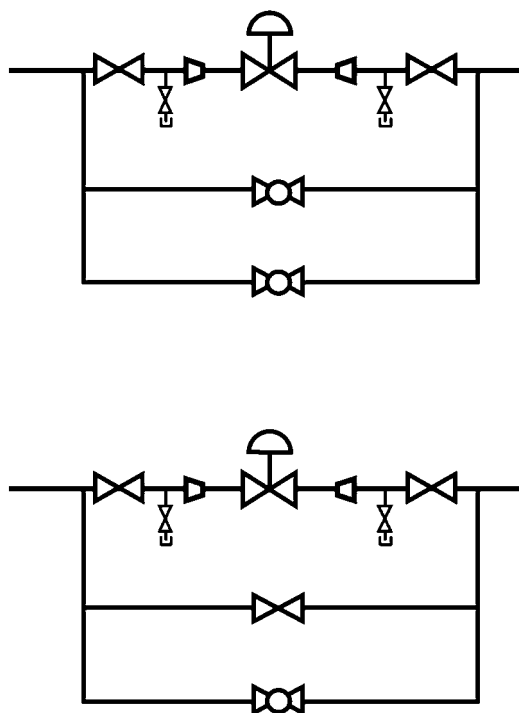


Figure 7.21 Options for dealing with requirement of large manual throttling valve.

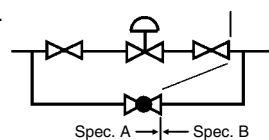
bypass valve for control valves in pressure control loop. Their logic is that essentially it is very difficult for an operator to adjust the pressure by a manual throttling valve. This is especially true when the system downstream of the control valve is designed based a lower pressure and we expect the operator to not only adjust the pressure but also to take care of the integrity of the downstream pressure!

Sometimes (and not always) a control valve decreases the pressure of the fluid severely such that the pressure of fluid on the downstream side of the control valve is very low. In such cases, it is a good idea to use less strong pipe spec on the downstream of the control valve. If the control valve decreases the pressure a lot and it is decided to change the spec of the pipe, spec break is needed on the control station. In such cases the spec break border should be placed on the downstream of the bypass valve and downstream of the isolation valve.

However, there could be less common cases that the condition on the downstream of the control valve is more aggressive even though the pressure is lower. For example, there could be an injected acid on the downstream. In such cases, based on the rule that says “the valves internals should be covered by the more severe piping material spec.,” the downstream piping material spec. should cover all the valves.

These two arrangements are shown in Figure 7.22.

Spec. A: severe spec.



Spec. B: severe spec.

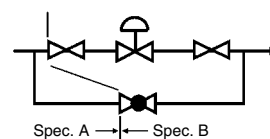


Figure 7.22 Spec. break on control valve station.

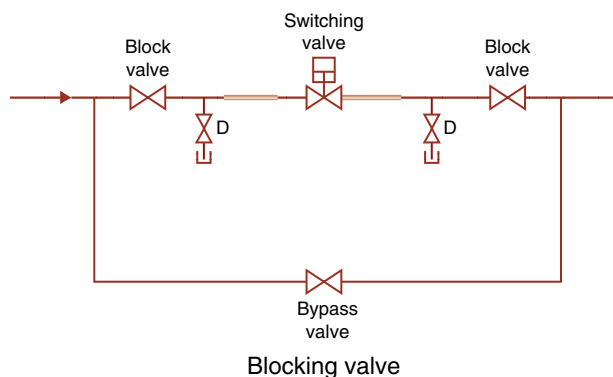


Figure 7.23 Details of a switching valve station.

7.13.2 Operating in the Absence of Switching Valves

We do not always see “switching valve station” similar to what we generally see in control valves as “control valve station.” There are some reasons for that. If a switching valve is functioning as a shutoff valve, it is very important for the valve to be closed and to stop flow when needed. A switching valve with a bypass manual valve provides the chance that the fluid still flows through the inadvertently left open bypass valve when it is supposed to be closed. That is the reason why companies are very cautious about implementing “switching valve station.”

However, if it is decided to implement a switching valve station, it would be similar to the one shown in Figure 7.23.

The main difference between control valve station and switching valve station is that in the latter, the bypass valve should be a blocking valve (and not a throttling valve).

7.14 Valves in Role of Unit Operation

Throttling valves create shear on the stream flowing through them. This means throttling valve could be used as energy injector to the streams. The P&ID representation

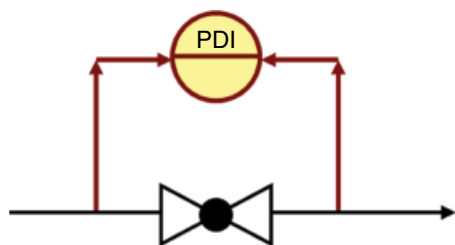


Figure 7.24 A valve to shear the fluid.

of valve in the role of unit operation can be seen in Figure 7.24.

In this role the amount of exerted shear in a function of closeness of valve or the pressure difference that it creates. Therefore a pressure differential indicator is required. In the more critical applications the manual valve could be replaced with a control valve, which could be connected to a process parameter. Such arrangement can be used for mixing (or a mixing valve), emulsifying, or viscosity breaking or other applications.

7.15 Special Valves

A valve that does not take any order from an external independent source is called special valve. A large effort in the valve industry has been made to develop new special valves for various applications. As “external independent source” (e.g. instrument air) is not always available, using special valves is very beneficial. Such lack of resources exists in remote areas as well as in small facilities.

In this section we will cover check valves, regulators, and safety-related valves.

A check valve is the first special valve to discuss as it is arguably the most common special valve. The applications of check valves were discussed before in Chapter 6.

Regulators are the other group of special valves. Their concept will be discussed in more detail in Chapter 13.

7.15.1 Check Valves

Check valves are arguably the most important special valves. They are also known as “non-return valves” and “flow direction assurance valves.”

These valves are the type of valves that do not receive any orders from an external source: they will be open or closed based on the desired direction of flow in process.

These valves also need an internal movable member to move away in the direction of flow and allow the flow in one direction and, when/if flow starts to reverse, move on the flow and prevent the reverse flow. There are principally three ways to implement a movable member: hanging, lifting, and flexible-shaped member. Each of these three could be designed in different structures, but the main structures are based on flapper, piston or ball, or an elastomer. Therefore check valves can principally be classified in four types: swing type, piston or ball type, and duck bill type.

Table 7.18 shows the different types of check valves and their symbols.

Conventional piston and ball check valves should be installed only in vertical position. However, spring-loaded piston check valves and spring-loaded ball valves can be installed in every position.

“Silent check valve” or “non-slam check valve” does not necessarily refer to any specific valve. It only refers to a check valve in which the backflow preventer element does not get closed suddenly in reverse flow. A conventional “slam-type” check valve prematurely breaks whenever the check valve is used in systems with frequent and heavy backflows. Here “heavy” means the reverse movement of a large volume of fluid. The conventional inexpensive “slam-type” check valves are conventional swing check valves and conventional lift-type check valves. The spring-loaded float check valves are generally considered as “non-slam check valves.” Tilting disk check valve and dual-plate check valve are other types of non-slam check valve.

Table 7.18 P&ID symbol for check valves.

	Movable part	Basic type	P&ID symbol
Hanging type	Flapper	Swing	
Lifting/pushing type	Piston	Conventional piston	
	Ball	Conventional ball	
Flexible-shaped type	Elastomer	Duck bill	

7.15.2 Regulators

Regulators are basically a control loop in a casing, or in other words they are “fully mechanical control loops.” Some applications of regulator are discussed in Chapter 14.

A typical control loop includes sensor controller and a control valve. If we gather all these three elements and put it in one small casing, then a regulator is produced. The advantage of regulator over a control loop is that it has less items and requires less maintenance and inspection and it does not need any external source of energy like instrument air. As for each process parameter we can have control loop; therefore there are regulators for each process parameter. Regulators are available for pressure, temperature, flow rate, and temperature.

There is no available regulator for composition. Composition sensors or in more technical term processed analyzers are too complicated to be able to be lumped in one casing as a regulator. Therefore regulators available in the market are pressure regulator, flow regulator, temperature regulator, and level regulator.

However, for pressure, there are two types of regulators: pressure regulators and backpressure regulators. These two types of regulators basically resemble two types of pressure control loop: in the first type the pressure sensor is upstream of the control valve, and in the second type the pressure sensor is downstream of the

control valve. In a pressure regulator the regulator senses the downstream pressure, while in backpressure regulator the regulator takes order from the pressure in the upstream of the regulator.

A pressure regulator has an internal valve that adjusts the opening of the valve to maintain a specific pressure in the downstream or upstream of the regulator.

In a flow rate regulator there is an internal valve that will throttle to adjust the flow in the pipe. It is important to know that by installing a flow regulator in a pipe, the flow regulator cannot guarantee that the flow is always on a specific value and all the way in a pipe.

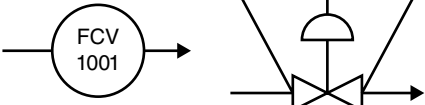

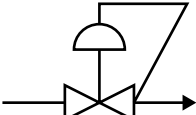

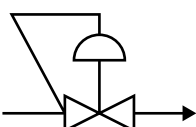
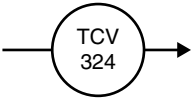

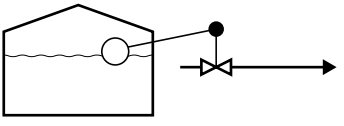

Inside of a level indicator, there is an internal valve in which the opening of the valve changes to adjust a specific perimeter line value for a liquid level in a container.

However, a temperature regulator has an internal valve that throttles to adjust the temperature in a pipe adjacent to the temperature regulator.

In P&IDs there are at least three footprints of regulators: their symbol, their tag, and their set point. The symbols for each type of regulators are shown in Table 7.19.

The regulators are tagged in the format of “*Process parameterCV*.” CV represents regulator. It is important to recognize the difference between tagging control valves and regulators. Control valves are tagged as “*Process parameterV*,” while regulators are tagged as “*Process parameterCV*.”

Table 7.19 P&ID symbol for regulators.

Process parameter		Corresponding regulator	Regulator tag
Flow			
Pressure	Downstream		
	Upstream		
Temperature			
Level			
Composition		Does not exist	Does not exist

It is very confusing that in tagging regulators CV represents control valve, but it is not the tag for control valve, but for regulator. “*Process parameter*” here can be any process parameter including “P” for pressure, “F” for flow, “L” for level, or “T” for temperature. The third footprint of regulator on P&ID is their set point. It is again important to recognize the difference between control loop and regulator here. On P&IDs, when there is a control loop, the set point of control loop is not shown on P&ID. The set point of control loops and P&IDs can be found in “Control and alarm set points table.” The set points of regulators are generally noted on P&IDs. Therefore, if there is a pressure regulator on a P&ID, the set point of the regulator should be mentioned beside the regulator on P&ID, for example, “set point; 20 kPag.” For a flow regulator the set point is mentioned beside the flow regulator symbol like this: “set point; 30 m³ h⁻¹.”

Regulators similar to control valves have failure position. They could be “FC” or “FO” or “FL.” However, the difference between regulator failure position and control valve failure position should be recognized. In control valves the designer has opportunity and the freedom to choose his/her favorite failure position. But this is not the case for regulators. Regulators have a natural built-in failure position that cannot be changed by the process engineer. For example, the failure position of all pressure regulators are FC.

The failure position of all backpressure regulators is FO and the failure position of flow regulators is always FL.

Figure 7.25 shows the set point and failure position of regulators on P&IDs.

The last thing about regulators is that they do not generally need bypass like what we see in control loops. As it was discussed we generally have a specific arrangement that we may need control loop station but there is no such thing for regulators.

As a rule of thumb, whenever we need to automatically control a process parameter, we need to put a control loop including sensor, controller, and control valve. However, in some cases, this system can be replaced by a

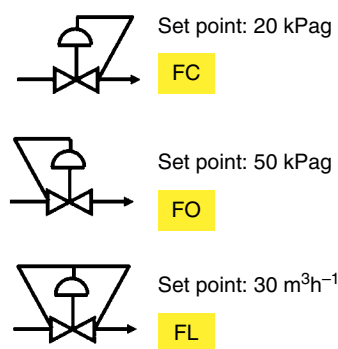


Figure 7.25 Regulator set point and failure position.

single regulator. Those are the cases in which we are seeking simplicity in the system and the service fluid is not dirty and the pipe size is not very big. Regulators are plugged easily if they are used in dirty services. Regulators also are not available in big sizes, say, larger than 6 in.

7.15.3 Safety-Related Valves

The safety-related valves are the valves that act when a process parameter wildly violates from its normal level. This violation could be in the lower or higher side of the parameter. Therefore unsafe condition happens when a process parameter goes to the higher side of the normal level or the lower side of that. Therefore it can be said that safety-related valves act when a process parameter goes much higher or much lower than its normal level.

As there are five different process parameters, namely, flow, pressure, temperature, level, and composition (Table 7.20), it can be assumed that there are at least five different safety-related groups of valves, which are roughly correct:

Pressure relief valves are the valves that open when pressure increases and reaches a preset value.

Vacuum relief valves are the valves that open when pressure decreases and reaches a preset value.

Temperature relief valves are the valves that open when temperature increases and reaches a preset value.

Table 7.20 Different groups of safety-related valves.

Process parameter		Corresponding safety valve	Action
Flow		Excess flow valve	Closes when flow goes beyond a preset value
Pressure	Internal	Pressure relief valve	Opens when pressure goes beyond a preset value
	External	Vacuum relief valve	Opens when vacuum goes beyond a preset value
Temperature		Temperature relief valve	Opens when temperature goes beyond a preset value
Level		No valve, only overflow nozzle	“Opens” when level goes high
Composition		Not available	Not available



Figure 7.26 Excess flow valve P&ID symbol.

These valves have a fusible plug that melts and ruptures in a specific temperature. This plug rupturing opens the valve.

Excess flow valves are the valves that stop the flow when the flow exceeds beyond the preset value. These valves are very common in the LPG industry as a safety measure. If there is a rupture in the pipe, the flow suddenly accelerates and goes to a flow rate higher than the preset value of excess flow valve. A typical symbol of excess flow valve is shown in Figure 7.26.

Some companies consider conventional check valves as “safety flow valve” and tag them as “FSV.”

Pressure/vacuum safety valves are very important in process plants and later will be discussed in more detail in Chapter 12.

7.16 Valve Combinations

Developing new valve combinations is another area in the valve industry. There are dozens of manufacturers that invent new valve combinations to satisfy a requirement in the industry. They combine several valves together and put them in one casing for the purpose of lowering the cost and making life easier for plant operators. Examples of such combinations are “flow preventer device,” “double block-and-bleed

device,” “gate valve with internal bypass valve,” and “minimum flow device.”

The downside of such devices is their lower redundancy. Because a fairly complicated system is placed in a casing, they are more prone to failure, and also failure may impact larger area. For these reasons the large companies that have enough budget generally prefer to buy individual items and assemble them in the plant rather than buying combined devices.

Table 7.21 shows few examples of valve combinations and the proposed P&ID symbol for them.

7.17 End of Valve Arrangements

There are different ways to “blind” a valve at the end of a pipe. To find out an appropriate way, the piping material spec. should be consulted. There are however rules of thumbs that are mentioned in Table 7.22.

Table 7.22 Arrangements of valves in end of pipes.

Pipe size	Joint name	P&ID schematic
NPS < 2"	Screwed plug	
	Screwed cap	
NPS > 2"	Blind flange	

Table 7.21 P&ID symbol of some valve combinations.

Combined system name	Schematic	P&ID symbol
Flow preventer device		
Double block-and-bleed device		
Gate valve with bypass valve		

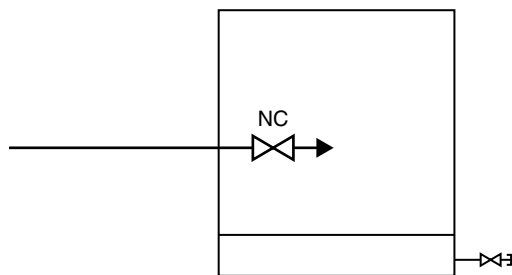


Figure 7.27 Drip catching system.

If a pipe ends to a valve or quick connecting fitting and needs to be opened frequently, it may need additional provisions to protect its environment.

Drip catching system may be needed wherever there is a non-coupled valve or quick connection in liquid services.

With today's ever-tightening environmental standards, the codes for allowable amount of liquid spills have become tighter and tighter. The liquid spill codes in different countries limit the amount of spilled liquid in a specific time period. Such limitation could be different for different liquids.

If spills are not allowable, a drip catching system should be installed at non-coupled valve or quick connection.

Figure 7.27 shows a simple drip catching system.

7.18 Valve Sizing Rule of Thumbs

It is important to know there is difference between valve body size and valve internal size.

Valve body size represents the size of its flanges or the size of pipe ends in order for a valve to be fitted to them without the need of any reducer/enlarger.

Valve internal size is the nominal size of the valve internal represented in the form of a pipe equivalent size.

Each valve body size may have up to three different valve internal sizes. What is seen on P&ID is only the valve body size. However, if a valve internal size is important, it should be covered in the notes area.

The rule of thumb for valve sizing could be split into two main types of valves: blocking and throttling.

The blocking valves generally have the same size that of the pipe. For example, a 6" pipe size needs a gate valve of 6", or a 10" pipe needs a 10" (gate) switching valve.

However, the case for throttling valves is not that simple. The size of throttling valves, either manual or automatic, needs to be specified during the design stage of the projects.

While this is true, as the manual throttling valves generally play role in very noncritical situations, the designers may decide not to do the specified size and

simply pick the same size that of the pipe for the manual throttling valve. It means that in noncritical service a 8" pipe may need a 8" globe valve.

However, this is not case for automatic throttling valves or control valves. While sizing of the control valve needs extensive calculation, we can just check the body size of control valves.

As a rule of thumb, the size of a control valve could be between minimum half size of the pipe and the same size of the pipe. The typical control valve size is one size smaller than the pipe size. It means that for a 10" pipe, it is not odd to see a control valve of 6" (half of 10") or even 10", but the control valve size on this pipe is typically 8" (one size smaller than 10").

This rule of thumb has exceptions especially for cases where pipes for some reasons are oversized.

The control valves on small bore pipes (say, less than 4") are generally the same size of the pipe.

7.19 Merging Valves

Valves are not generally merged. In Chapter 6 we discussed about the cases wherein a pipe can be merged. Whenever a pipe is merged and handles more than one duty, all the valves on the pipe, including manual and automatic ones, are shared as well.

7.20 Wrapping Up: Addressing Requirements of Valve During the Life Span

In this portion we check our design to make sure we cover all the needs of valves during each phase of plant life. As it was discussed before, these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item:

- 1) Normal operation of valves: The required considerations are already covered.
- 2) Nonroutine operation (reduced capacity operation, start-up operation, upset operation, planned shutdown, emergency shutdown): This phase of plant life needs two components to be handled: process component and control/instrumentation component. The process component is discussed here, but the control/instrumentation component will be discussed in Chapters 13–15.

Generally speaking, valves do not need much consideration during the P&ID development to cover nonroutine operations. These valves are very robust items that can handle different conditions easily.

On the other side automatic valves need some attention. We already talked about control valves.

- 3) Inspection and maintenance: General consideration regarding inspection and maintenance of all items will be covered in Chapter 8. Here, however we cover the specific requirements.

Manual valves are very robust items and do not need much attention during the operation.

On the other side automatic valves needs some provisions. We already talked about the required provisions for control valves and switching valves. One other inspection for switching valves is “stroke test,” which does not have any footprint on P&IDs.

- 4) Running plant in the absence of a valve: This stage was also covered before.

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- 2 “Two Pressure-Reducing Valves In Series – Xylem Applied Water Systems – United States”. Xylem Applied

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8

Provisions for Ease of Maintenance

8.1 Introduction

P&ID development for the purpose of inspection and maintenance was briefly discussed in Chapter 5. This concept, however, is so important it needs a chapter devoted to it.

This chapter is about the required provisions that should be considered in P&IDs to facilitate an inspection and/or maintenance operation.

These provisions are used for the purpose of facilitating inspection and maintenance, and also sometimes as part of a shutdown system. The instrumentation requirements of safe shutdown are mentioned in Chapter 15 but here the process requirements are mentioned.

In this chapter we discuss the requirements of three elements of each process plant: equipment, utilities and instruments regarding ease of maintenance and inspection.

The inspection/maintenance of instruments and utility systems is also mentioned here.

8.2 Different Types of Equipment Care

From a maintenance viewpoint, “equipment care” can be categorized based on the location of applying care and the timing of the care.

The care could be done in workshop, or “in-workshop care,” or could be done when the piece of equipment is not dislocated, “in-place care.”

The in-place care can be categorized further into “in-line” or “off-line” operation.

The in-line care is performed during the operation of the unit of interest while in off-line care the unit of interest has to be pulled out of operation for the required care. We love equipment and units that need in-line care rather than off-line care. One attempt of equipment fabricators is to develop new equipment with less need for off-line care and more in-line care. For example in

upstream oil extraction there is a piece of equipment called an FWKO drum or “free water knock out drum.” This vessel may see heavy settlement of sands in it. In older days, at specific intervals the FWKO drum would be taken out of service for “de-sanding operation” but these days there is equipment with automatic de-sanding systems that can remove the sand from them during the normal operation of FWKO drums.

Although it could be said that major maintenances occurs in workshops and only minor repair can be done in-place but it is not the cases always. There are some huge equipment that all of their maintenance-small or big- should be done in field, or “in-place care”. Tanks are a famous example of them.

Each of these two different types of equipment care forces us to provide different types of provisions for equipment on P&IDs.

Table 8.1 shows the responsibility of different groups in process plants for different types of equipment care.

8.3 In-place In-line Equipment Care

“In-place in-line care” means any activity that takes care of a process item without any disturbance to the normal operation of the element or with just mild disturbance.

There are mainly two types of activities in this group: activities related to the rejuvenation of the equipment and activities related to monitoring the equipment.

Rejuvenation activities are needed for equipment that is inherently in an intermittent or semi-continuous mode of operation. One example of rejuvenation activities has already been stated as de-sanding in FWKO drums.

The rejuvenation activities could be triggered by one or a combination of these parameters. They can be initiated by receiving a specific sensor signal (event based), or they could be initiated after a specific time interval (time based), or by the decision of the operator.

From the type of operation viewpoint, a rejuvenation process can be done manually or automatically. In manual

Table 8.1 Different types of equipment care.

	In-line	Off-line
In-place	By process group	By mechanical group
In-workshop	Not applicable	By mechanical group

rejuvenation all the operation are performed by operator(s) in the field. In automatic rejuvenation the operation can be done fully automatically without any action by an operator, or can be initiated and observed by an operator from the control room (and not the field).

Manual rejuvenation is done if the rejuvenation doesn't need to be very frequent, or the rejuvenation steps are not very complicated, or if the unit of interest doesn't have very critical role in the plant. Otherwise automatic rejuvenation should be performed.

Below are a few examples of equipment requiring rejuvenation activities.

Sand filtering in water treatment operation is generally a semi-continuous operation. After a certain time period, the filter should be taken out of operation for backwashing. This type of operation should be implemented during the design and be shown on a P&ID. The backwashing could be manual (in small systems) or automatic. The backwashing operation is a type of "in-place in-line care."

Another example is a soot blower for heating coils in the convection section of fire heaters and boilers. These coils get covered with soot after a while. To keep the same efficiency of the fire heater or boiler this soot should be removed from them. A soot blower is a moving set of perforated pipes that slide into the convection section of fire heaters and, by injecting steam, dislodge the precipitated coke or soot.

Another example is the cleaning of bin filters. Bin filters are a type of in situ treatment of bin vents. Bin and silo vents may have a large amount of dust. To remove the dust from the outgoing gas (air) a bin filter can be installed on the vent of bins or silos. It is very common to use bag filters in the role of bin filter. In each bag filter there are several filter bags (socks) that separate the dust from the gas stream. These socks get full after a while then need to be rejuvenated. The rejuvenation of each sock can be done by backflowing an air jet in it. As there are several socks in each bag filter. One sock can be cleaned during normal operation with marginal impact on the operation of the bag filter.

The last example of rejuvenation is the regeneration of ion exchange systems. Ion exchangers in, for example, water treatment systems don't work on a fully continuous basis. They function to adsorb the ions on their "exhaustion period." At the beginning of this period the resin beads inside of the ion exchange vessel are fresh and ready to adsorb ions. After a while the resin beads are "filled"

with ions and become exhausted. To return them back to their fresh condition, the resin beads should be "regenerated." This regeneration, which is a type of rejuvenation, is done by sending a solution through the ion exchangers.

A large group of rejuvenation operations are categorized as "cleaning." The different cleaning methods are discussed later in this chapter. It is very common to do automatic rejuvenation. The automatic rejuvenation can be provided by a mechanical unit that recirculates the cleaning agent or object. The automatic rejuvenation systems are any of two main types of "mechanical-in-place" (MIP) systems, or "clean-in-place" (CIP) systems. These systems will be discussed later in this chapter.

Activities related to monitoring equipment in in-line care are mainly on the shoulders of the "rounding" operator. The operator could be equipped (beyond just personal safety equipment) or non-equipped. If the operator is non-equipped, he/she relies on his/her five senses. Among them, the taste sense never is used. Below is a non-inclusive list of the use of the remaining four senses of the rounding operator:

- Vision: leakage, vibration, overflow of tanks, checking levels, checking flame color and shape
- Hearing: vibration, cavitation, hammering, PSV release, explosion
- Touch: vibration
- Smell: fire, leakage, PSV release to atmosphere.

To facilitate the duties of the rounding operator, the process engineer can put some items on the P&ID. The examples are sight glasses to check liquid levels, catalyst levels, or filtering media levels, and peep holes to check the color and shape of flames in a furnace or boiler.

An equipped rounding operator can have some small measuring devices to check some parameters that cannot be accurately checked without. It could be a portable pressure gauge, portable temperature sensor, etc. If it is the plan to have an equipped rounding operator, the process engineer needs to provide "test points" on important points of the system for him. They could be a "pressure tap" (PT), or "temperature point" (TP). However, these points should not be very important parameters because if one parameter is very important to check and monitor, it should be part of a control loop. An example of a PT could be the suction side of centrifugal pumps. In some companies the acronym of PP is used for "pressure point" and TW for "thermowell." This concept will be discussed in more detail in Chapter 13.

8.4 In-place Off-line Equipment Care

"In-place off-line" care could be for small repair jobs, inspection, or as a preliminary step for "in-workshop off-line care" activities.

It generally starts with de-energizing the equipment. Then the piece of equipment should be isolated from the rest of the plant and tagged to show it is under maintenance. Such activity generally is known as LOTO in plants, representing “locking out–tagging out.”

The next step is venting and draining the system and finally some sort of cleaning.

Such cleaning may include chemical/solvent cleaning, steaming-out, pigging operation, etc. Depending on the type of required in-place off-line operation, different items should be implemented (such as chemical cleaning valves). However, a set of items to provide a positive isolation is usually a must. This “set” comprises isolation valves, drain and vent valves, etc.

The detail of operation will be discussed in section 8.6.

8.5 In-workshop Off-line Equipment Care

“In-workshop off-line” is generally for fundamental maintenance.

The requirements are similar to “in-place off-line” care in addition to a last provision for the ease of equipment removal.

This means it starts with LOTO, then venting and draining, and then cleaning. The last step considers provisions to allow the equipment to be removed from its foundation easily and safely. However, all the requirements are not always shown on P&IDs. For example, if equipment needs to be hoisted for removal, we usually do not show the hoist on the P&ID.

The detail of operation will be discussed in section 8.6.

8.6 Preparing Equipment for Off-line Care

Off-line care can be done in-place or in-workshop. Both of them need the same set of preparations with one exception. For in-workshop care an additional step at the end is needed to “untie” the item of interest from the tangled piping.

To prepare a piece of equipment for off-line care these steps should be taken:

- 1) Isolation of the equipment of interest
- 2) Bringing the equipment of interest to non-harmful condition
- 3) Making enough room for equipment removal IF the goal is off-line in-workshop care.

It can be seen the last step is applicable only if the equipment needs to be transferred to the workshop.

In the following sections these three steps are expanded.

8.7 Isolation

When deciding on isolation systems, three questions need to be answered:

- 1) For what equipment should it be added?
- 2) Which type of isolation system should be used?
- 3) Where does it need to be placed “around” the equipment?

8.7.1 Requirement of an Isolation System

The first question that needs to be answered regarding the need for an “isolation system,” for a piece of equipment or a portion of a plant, is whether it is necessary to do something on that piece while the rest of the plant (or unit) is operating or not.

To answer this question, two parameters must be compared with each other: MTBF (mean time between failures), and “time interval between turn-arounds.” MTBF was introduced in Chapter 5.

For example, a company may state that the MTBF for control valves in their refineries in the Middle East is six months. This means that, based on their experience, a typical control valve will break down every six months and will need to be repaired.

Turn-around is the practice of shutting down the whole plant for a few months to do complete servicing and maintenance on all of its equipment.

The time interval between turn-arounds used to be one or two years, but now companies try to push it to four to six years.

Now the answer to the question is: if the MTBF of a piece of equipment is shorter than the time interval between turn-arounds, then there is a need for placing an isolation system around the equipment.

It is obvious that if a piece of equipment will break down every three years, but the whole plant is planned to be shut down for turn-around every two years, there is no need for an isolation system. As was mentioned, isolation is needed “if a piece of equipment needs repair while the rest of plant is operating.”

Therefore providing isolation valves is not necessary for all the equipment in a plant. Isolation valves are required to isolate the equipment from the rest of plant if we know this equipment needs “off-line care” in a period shorter than the plant turn-around period.

For example if you have a switching valve that needs off-line care every three years – based on your experience and historical data – and the plant you are developing the P&ID for it has a planned turn-around every two years, there is no need (at least theoretically) to put isolation valves upstream and downstream of it.

This concept is shown in Figure 8.1.

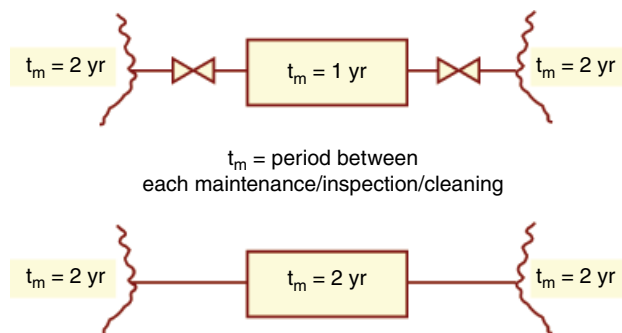


Figure 8.1 Dependency of need or lack of need for isolation systems for items.

However, we know (based on the concepts mentioned in Chapter 5) that when a process item is taken out of operation, consideration should be made of the way the plant (or unit) should operate in the absence of that element. There are some cases where there is no technically acceptable solution that effectively mitigates the lack of item so that the rest of plant can operate with minimum impact. In such cases one may essentially challenge the need for placing an isolation system for that specific item. One example is heat exchangers. When you put a heat exchanger on a stream in your plant you want to increase or decrease the temperature (and maybe the phase) of your stream. Most likely that temperate change is so important that you were forced to put in a heat exchanger. Now, what would you do in the absence of a heat exchanger? If you review the available options mentioned in Chapter 5, you may see it is very hard to find an attractive solution. If this is the case, why you should bother to put in an isolation system “for the time you need to pull the heat exchanger out of operation”?

Therefore the other question you need to ask yourself before placing an isolation system is whether you can “afford” to be without that piece of equipment in the plant or not?

This is the reason that some companies don’t provide isolation systems for their heat exchangers apart from a few exceptions.

Exceptions could be when there are spare heat exchangers available in parallel, when there is an automatic cleaning system for the heat exchanger that can clean it in a short time, and/or when the target stream leaving the heat exchanger goes into a large container. In the first case we obviously need an isolation system to bring the spare heat exchanger into service. In

the second and third cases, we again need an isolation system, but we do nothing during the time we are lacking the heat exchanger; however, we still can afford it because it is a short time and/or the disrupted stream (with non-suitable temperature) goes into a large container and is thermally equalized with the bulk fluid in it in a way that only a small and acceptable temperature change can be observed.

8.7.2 Type of Isolation System

The second question is: what is the isolation arrangement? To answer this question, the concept of isolation should be discussed.

Isolation in this context means the segregation of a piece of equipment, or even a portion of the plant, from the rest of the plant while the plant is operating. Isolation is done using an “isolation system.”

The general concept of an isolation system is shown in Figure 8.2.

The purpose of isolation could be inspection, cleaning, in-place repair, workshop maintenance, etc.

One may say that isolation can be provided simply by closing the inlet and outlet valves. You can see such an arrangement for a pressure gauge in Figure 8.3.

The valve symbol shows a valve that we generally call a “root valve” because it is installed at the root of the pressure gauge. This root valve serves as an “isolation valve” and is a ball valve or gate valve. If someone needs to inspect and/or re-calibrate the pressure gauge, they can do it without emptying the vessel.

Although a single simple valve can be accepted as an isolation system in some non-complicated systems (like instruments), it is generally not accepted for the isolation of equipment.

“Root valves” are a type of isolation method used for instruments.

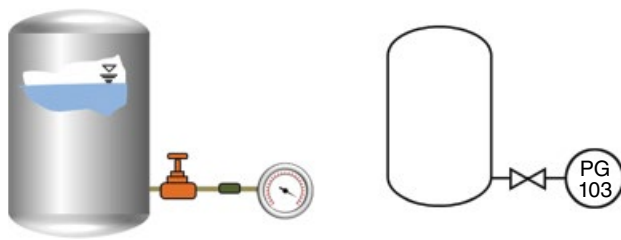


Figure 8.3 Root valve for isolation of a pressure gauge.

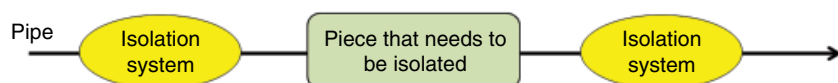


Figure 8.2 General overview of isolation.

In utility networks root valves are also common. In utility networks root valves are “planted” in suitable locations (e.g., at the root of each sub-header or important sub-headers) to make sure a broken part of a utility network doesn’t force us to shut down the whole of the network of that utility.

There are basically three types of isolation arrangements:

- 1) Block valve and blind
- 2) Double block valve (and blind) and bleed (DB&B)
- 3) Block valve (and blind) and removable spool.

The type of isolation is determined by process engineers based on client guidelines. There is always the chance of accidentally jeopardizing the effectiveness of the isolation, which was put in place to isolate a piece of equipment for the purpose of inspection or maintenance, so that fluid flows toward the equipment which was supposed to be free of any. If an isolation arrangement provides more opportunity for this potential jeopardy, it is named a weak isolation method, and if it blocks the majority of potential jeopardy, it is named a strong or positive isolation arrangement.

Out of the three mentioned isolation systems the first one is the weakest type of isolation and the last one is the most positive type of isolation.



Figure 8.4 Block valve and blind.

The first type of isolation system is shown in Figure 8.4. The P&ID symbolism of it is shown in the second row of Table 8.2.

Table 8.2 shows different types of blinding arrangements.

The second type of isolation or DB&B is the type that, on the top of the first isolation valve and blind, there is another valve that should be closed for the purpose of isolation. Between these two isolation valves there is one drain valve (bleed valve) that should be opened and monitored by an operator. If there is any even small fluid discharge from the bleed valve, it means there is a problem in the DB&B and inspection or maintenance should be stopped.

The third type of isolation is the type where again there are two block valves in series AND the piece of pipe in between is removed for the purpose of isolation.

Table 8.2 Different types of isolation.

Type	Symbol	Positiveness
1 Block valve (with or without lock)		Generally not acceptable as a positive isolation method for voluminous process items
2 Block valve (with lock) and blind		
3 Double block valve (with lock) and bleed		
4 Block valve (with lock) and blind and removable spool		

More safe isolation

You probably realized that “blinds” are the vital component of the different types of isolations. There are at least two types of blinds: spectacle blinds and spade blinds.

Figure 8.5 shows a spectacle blind when it is in use. In the figure the pipe is obviously NOT blinded since the solid portion of the blind is up.

Table 8.3 summarizes the different features of these blinds.

It should be noted that in Table 8.3 the position of the blinds in the open and close positions are shown but in P&ID we only show the “open” position of blinds.

A decision needs to be made about the right type of isolation method for each application. The isolation method depends on the type of item that needs to be isolated, the fluid type, and the pressure of the fluid in the item.

The first parameter is the type of process element. The process element could be so small that even the hands of

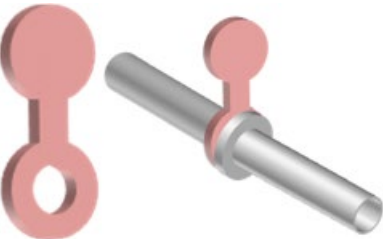


Figure 8.5 Spectacle blind: outside of the pipe and in the pipe.

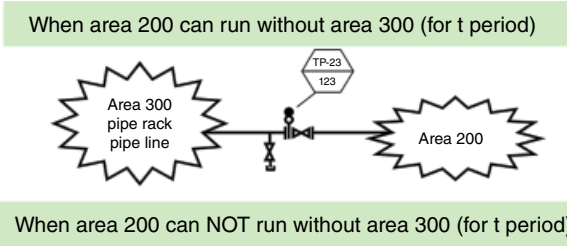


Figure 8.6 Isolation for utility streams across different areas.

the operator cannot get into it. Such items need the least positive isolation. If the item is head-in or worse, walk-in, it needs a stronger isolation system. Walk-in equipment is a piece of equipment that is large enough that a person can walk in it, but it is not designed for human to live in it. The technical name for such spaces is “confined space.”

“Confined space” is a concept that has a technical meaning and in different jurisdictions also has a legal meaning.

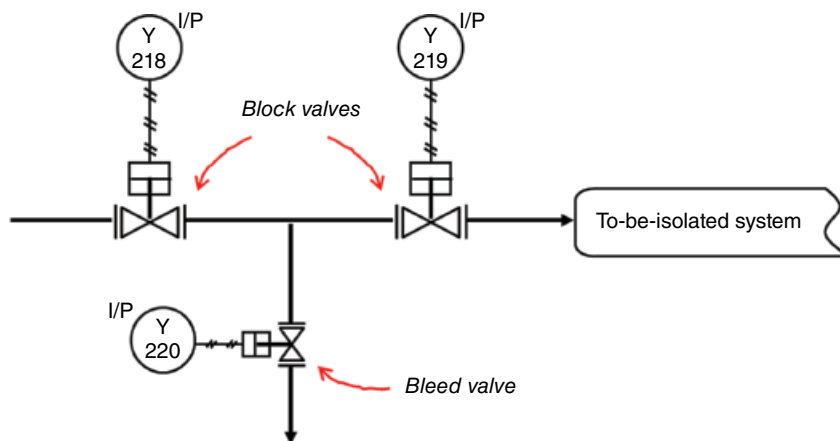
A utility network or a pressure gauge can be isolated with a simple root valve as they are not voluminous.

The second parameter is fluid type. If the fluid is non-innocent the more positive isolation system should be used. In technical terms, fluids like steam, acids, and toxic materials need a strong isolation system.

Table 8.3 Features of different types of blinds.

	Spectacle blind (see Figure 8.8)	Spade blind (spacer)
Real shape		Open Close
P&ID symbol	Open Close	Open Close
Pros	It can be seen by the operator whether a pipe is blinded or not	Not heavy; easy to handle
Cons	Sometimes too heavy, especially for large pipe sizes with large wall thickness	The operator cannot see whether the pipe is blinded or not; easy to misplace the mate
Applications	Only for smaller pipe sizes, typically less than 12"	Generally for larger pipe sizes, typically larger than 10"

Figure 8.7 Automatic double block and bleed system.



The other parameter is the pressure of the fluid. If the pressure is higher we may need a more positive isolation system. For example isolation of high pressure steam should be stronger than low pressure steam.

The other location that we generally put an isolation system in is when a utility stream goes from one area to another. The type of isolation system depends on the fluid type and pressure, and on the border of the areas, unless two areas are so interrelated that one cannot be run when other one is shut down. This concept is shown in Figure 8.6 for a high pressure steam stream.

Isolation may also be needed as part of the safe shut-down of a unit. In such cases the isolation is done fully automatically and probably without usage of blinds (Figure 8.7).

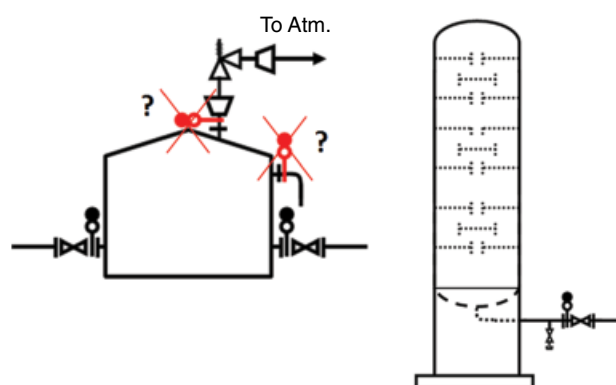


Figure 8.8 Location of isolation systems.

8.7.3 Placement of an Isolation System

The answer to the third question is that isolation system should be added to all downstream and upstream connecting pipes, and as close as possible to the equipment.

Some companies, however, challenge this, and question whether it is a really necessary to put an isolation valve on a pipe that goes to atmosphere (Figure 8.8).

8.7.4 Inbound Versus Outbound Blind Location

The question arises when talking about the blind is its location with respect to the to-be-isolated system. The blind (either of spectacle type or spade type) should be placed on the isolating valve but the question is whether

it should be placed closer to the to-be-isolated equipment, inbound, or away from the to-be-isolated system, outbound (Figure 8.9).

The difference is that in outbound blinding the to-be-isolated system doesn't need to be emptied before blinding (Table 8.4).

Some companies prefer outbound blinding because there is more flexibility in operation, but some other companies prefer inbound blinding because it allows "the only correct isolating sequence."

8.7.5 Merging Isolation Valves

There is one opportunity for saving isolation valves. This is a good cost saving strategy if pipes are of large size, e.g. larger than 12". This concept was covered in Chapter 7.

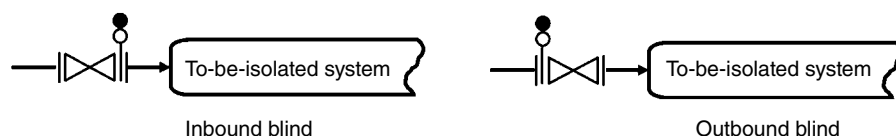


Figure 8.9 Inbound versus outbound blind location.

Table 8.4 Sequence of isolating for different positions of blind.

Blind position	Sequence of isolating
Inbound blinding	Closing valves—emptying—blinding
Outbound blinding	Closing valves—emptying—blinding or Closing valves—blinding—emptying

Table 8.5 Actions needed to bring the process parameters into the safe range.

Process parameter	Action needed
1 Flow rate	Already done through isolation
2 Temperature	Takes time
3 Pressure	Emptying—venting
4 Level (liquid)	Emptying—draining
5 Composition	Cleaning

8.8 Bringing the Equipment to a Non-harmful Condition

The next step for making equipment ready for inspection/maintenance is bringing it to a “non-harmful condition.” It means all the five process parameters should be in a safe range. To bring each parameter within the safe range the actions below should be taken:

- Making flow rate safe: when the equipment is already isolated from the plant, this is not applicable.
- Making temperature safe: allowing time lapses, for cooling down (or warming up) streams.
- Making pressure safe: gas/vapor should be displaced. Emptying and then venting may be needed.
- Making levels safe: liquid should be displaced. Emptying and then draining is needed. It is applicable for every pipe and every voluminous item in liquid services, from tanks and vessels to pump casings.

The meaning of “emptying” here is removing gas or liquid from the process item using existing equipment and in large flow rates. This concept will be discussed later in this chapter.

- Making compositions safe: this is especially important for equipment that will be in contact with operator’s hands and body or is in confined spaces and a safe breathing atmosphere is needed. Cleaning of the body and the internal atmosphere of the equipment should be carried out to make it composition safe.

The actions that should be taken to bring each process parameter into the safe range are summarized in Table 8.5.

Based on the above discussion, to make a piece of equipment safe to handle three steps should be taken:

- 1) Cooling down (or warming up for cryogenic systems)
- 2) Emptying and then draining/venting
- 3) Cleaning/purging.

These three actions are discussed below.

8.8.1 Cooling Down

The first step for making equipment safe to touch is making the temperature safe. The temperature of the equipment of interest could be excessively high or excessively low and not safe to be handled by personnel. Such temperatures should be brought to a safe temperature, which could be loosely defined as ambient temperature. For this activity we generally don’t do anything other than leaving the isolated equipment for few hours or days to come to equilibrium with ambient temperature.

For items that are in contact with cooling water during their normal operation, the cooling water circulation may be used to accelerate the cooling down of hot items.

In process items that work in gas phase, sometimes the cooling down is accelerated by sending them through a cold neutral gas (like nitrogen).

8.8.2 Emptying and Then Draining/Venting

For different purposes, equipment, tanks, vessels, etc. should be emptied. Emptying could be for the purpose of inspection and maintenance, or shutting down a plant, a unit or a plant. Emptying means removing all the internal process material from the equipment or containers. The process material could be in the form of solids, liquid, or gas. Removal of each of these materials requires a specific method.

The most preferable way of removing processed material from a piece of equipment is removal using an existing system. For example, to remove a liquid out of a tank the best solution is to remove the liquid with one of the existing pumps, which is already connected to the tank, and this discharges the liquid out of the tank.

However, this solution is not always available and sometimes the equipment available cannot empty the piece of equipment completely. For example, if a pump is used to remove a liquid from a tank the pump can reduce the liquid level in the tank to a low liquid level or low low liquid level, depending on the design. The reason for this limitation is that if the liquid level drops lower than a certain number the pump will cavitate. Therefore, we need to seek other solutions – as a complementary solution or as a sole solution – to empty the piece of equipment.

This solution is to remove the fluid from the equipment by gravity. In this solution we need to design the system in such a way to be able to remove the fluid from a piece of equipment by fluid gravity and without using any external equipment or energy.

It could be questioned why we don't do this in the first place; gravity draining doesn't need any energy while the first solution needs equipment and also some external energy. The answer is that gravity draining is generally a very slow process. Removing fluids from equipment using their gravity usually takes a long time. Because of this, for emptying we prefer to use an existing pump in the first place and then rely on gravity draining.

To remove liquids, first of all the equipment or equipment internals should be oriented to naturally direct the liquid toward some drain valves to drain liquids and fully empty the piece of equipment.

Drain valves are usually small valves that the designer should put in different locations on piping and on equipment to provide drainability for the system. The drain valves are a type of on/off valve and could be gate type or ball type.

To remove gas and vapors from equipment, containers, and piping, we similarly need to put in some vent valves. Vent valves again are small valves in the form of gate or ball valves.

Figure 8.10 shows drain and vent valves on different equipment.

The responsibility of the P&ID developer is to decide on the location, number, and size of drain and vent valves in the system.

If there is some solid or semi-solid material inside of a piece of equipment, emptying is more difficult than for fluids. The P&ID developer needs to put adequate number of nozzles in the piece of equipment to be able to empty the equipment by gravity. In such cases the piece of equipment should have a high height to diameter ratio, otherwise the solids won't be removed from it by gravity. In difficult cases, machines (like a vacuum cleaner) may need to be used for removing solids or semi-solids from the equipment.

8.8.2.1 Location and Number of Drain/Vent Valves

The required number of drain/vent valves is decided based on several points.

The first point is that draining and venting of a piece of equipment or system should be done in a reasonable time. This gives a rule of thumb about the number of drain and vent valves that are needed to drain a specific portion of a

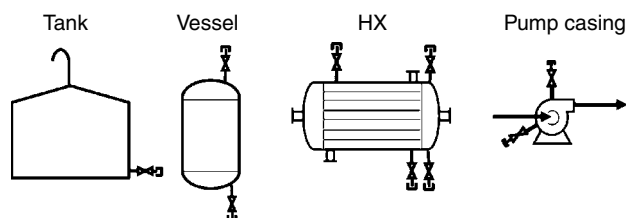


Figure 8.10 Vents and drains on different equipment.

process plant. For example, it is not acceptable that a small tank (less than 300 m³) be drained within three days!

The second point about placing drain and vent valves is that each drain or vent valve has a specific coverage area (vents can cover a bigger portion). This means a single drain valve on a piece of pipe cannot empty all the pipe from point A to point B. At least one drain valve should be placed for each piece of pipe between two isolation valves, or isolation/check valve, or isolation valve/safety valve, or isolation valve and concentric reducer, or enlargers.

Figure 8.11 shows the spans on pipes that need drain valves.

However, additional vent and drains may be needed if the pipe route goes up and down. The general rule is that at each summit point we need to put a vent valve, and in each valley point we need to put a drain valve. However, these high point vents and low point drains cannot be "seen" on a P&ID. They only can be recognized from a

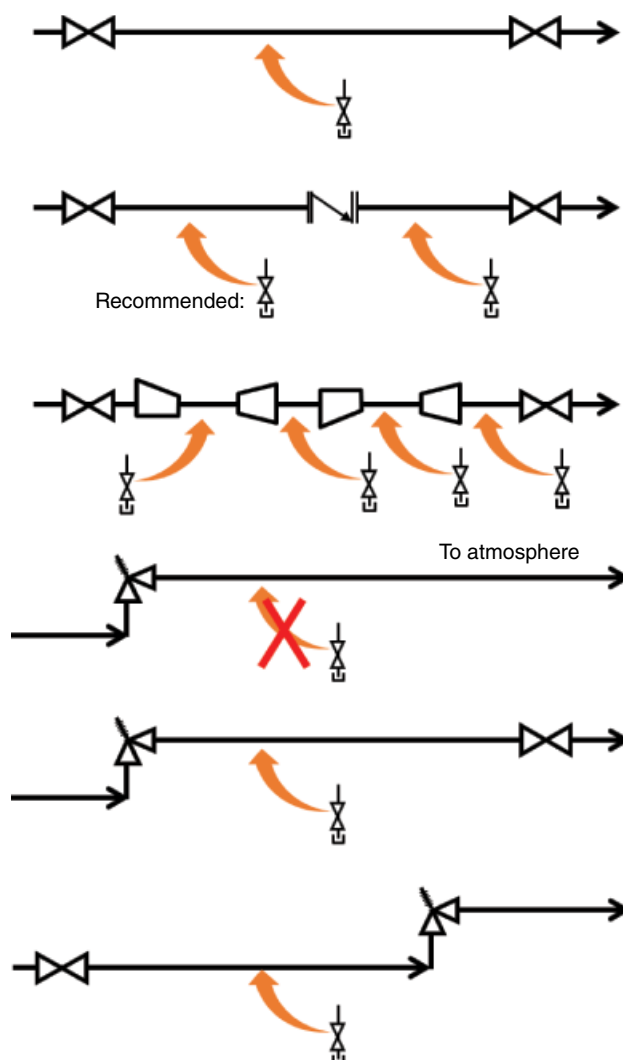


Figure 8.11 Requirements of drain valves on pipes.

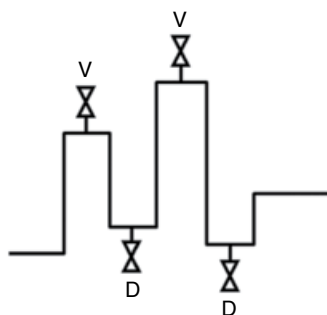


Figure 8.12 High point vents and low point drains.

piping model and therefore they should be placed on the P&ID after completing a piping model (Figure 8.12).

Some companies come back and show these high point vents and low point drains on their P&IDs but some others don't show them at all on P&IDs.

Drain valves are usually placed vertically and downward and on the bottom of items while vent valves are placed vertically and upward and on the top of items.

8.8.2.2 Size of Drain/Vent Valves

The smallest commonly used drain/vent valve is $\frac{3}{4}$ ". The largest size of drain/vent valve should be such that the flow out of the valve during the emptying procedure is not dangerous and is manageable. Generally speaking we limit the size of drain/vent valves to 2" unless the drain/vent valve is inside of a dyke.

For draining and venting pipes the company guidelines should be followed or the rule of thumb in Figure 8.13 can be used.

When the space that should be emptied is very small and mainly a flat system there is no need to have the dedicated drain valves and dedicated vent valves. For example, for pipes that are in the horizontal position and are smaller than 2", one valve can work as a drain and as a vent valve.

Some companies have guidelines about the selection of drain/vent valves. However, if there is no guideline a 1" drain/vent valve can be used. For vent valves the size could be one size smaller than drain valves because releasing gas or vapors from valves is easier than liquids. If a liquid has a viscosity higher than 50 cP or the electrical

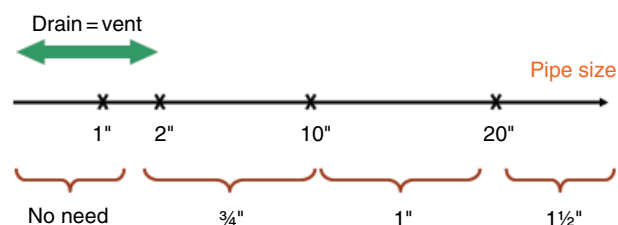


Figure 8.13 Rules of thumb for sizes of drains and vents for pipes.

Table 8.6 Approximate drain sizes required for voluminous equipment.

Container diameter (m)	Drain size (")
<2	1½
2–4	2
4–6	3
>6	4

conductivity of liquid is low the drain valve should be one size bigger.

To empty liquids from voluminous equipment we may need to put multiple drain valves. Generally speaking we need multiple drain valves to solve the issue of lack of draining in the cases where there is a chance of equipment foundation settlement. The other cases for which we may need multiple drain valves are for compartmented containers.

If the volume of the equipment is less than half a cubic meter, like pump casing, a single $\frac{3}{4}$ " drain is enough.

As a rule of thumb, to empty liquid from a container or voluminous equipment with a diameter of 2 m then a drain of 1½" should be used, and increase one size per each 2 m of diameter of container.

Based on this rule of thumb we can list the required drain size for voluminous equipment in Table 8.6.

For very low level horizontal vessels (when $L/D > 2$), the drain can be one size bigger for quicker liquid draining.

For very high vertical vessels (e.g. towers with a height of more than 15 m), the drain can be one size bigger to decrease the landing distance of the draining liquid.

Drains/vents may need to be routed to a safe location. The ultimate destination of the fluids out of vents and rain is discussed later in this chapter.

8.8.2.3 Other Usages of Drain/Vent Valves

Drains and vents can be used for other reasons too. For example vents are commonly used to de-aerate the equipment for start-ups too.

The other usage of drain and vent valves are for hydrotesting. However, sometimes another set of valves is placed for the purpose of installation. It is very common rule in companies that they don't show hydrotest valves on their P&IDs.

There are several valves attached to the pipes and equipment with the same size and same location. Some companies try to "merge" them together to save money. Examples of these valves are vent, drain, steaming out, chemical cleaning, etc.; however, such merging does not always provide a huge cost saving as these valves are generally tiny.

8.9 Cleaning

Making composition safe is basically removing process material left after simple venting and draining.

Making composition safe is generally a cleaning or purging operation. Cleaning could be a pre-requisite of sending a piece of equipment for maintenance or the sole reason for taking a piece of equipment out of operation.

However, “making composition safe” is especially important for equipment that needs walk-ins for inspection and/or maintenance.

Different methods of “making composition safe” for different phases of trace material inside of equipment are summarized in Table 8.7.

Cleaning could be needed for any of the reasons below:

- 1) To make the equipment ready for off-line in-place care
- 2) To make the equipment ready for off-line in-workshop care
- 3) As a stand-alone operation to return a unit to full efficiency and or minimizing corrosion
- 4) As part of shutdown system for safety purposes.

The cleaning is done through a “system” manually or automatically. If the goal of cleaning is supporting safety for a shutdown it should be fully automatic. Cleaning as a stand-alone operation could be automatic too.

The automatic cleaning could be in the form of MIP systems, or “CIP systems and other forms. “Water-in-place” (WIP) and “steam-in-place” (SIP) are examples of CIP.

8.9.1 Solid/Semi-Solid Removal Methods

Here we are talking about solid material inside of process elements that cannot be removed by liquid cleansing fluid. If they are easy enough to be removed by liquids, then this is less expensive. Liquid cleaning is discussed in the next section.

Table 8.7 Features of different cleaning/purging methods.

Type of “dirt”	“Dirt” removal method
1 Solid/semi-solid removal	<ul style="list-style-type: none"> • Manual • Machine-assisted (mechanical)
2 Liquid washing	<ul style="list-style-type: none"> • Flushing: using utility water • Steaming out: using utility steam • Chemical cleaning: using chemical solutions or solvents
3 Gas/vapor purging	By neutral gas: <ul style="list-style-type: none"> • Inserting • Ventilation

Table 8.8 Requirements of roddability and piggability.

Roddability	Piggability
<ul style="list-style-type: none"> • Full-port valves • Three-way instead of elbows • Sectionalization of pipe circuit (placing flange in specific intervals) • Etc. 	<ul style="list-style-type: none"> • Full-port valves • Long radius elbows (and not standard elbows) • Etc.

The solid material removal methods can be categorized as solid removal of pipes and solid removal of containers.

For pipes, the solid removal could be needed if they carry heavy precipitating fluids. In such cases rodding or pigging can be performed for solid removal.

In both rodding and pigging a solid object removes solids deposited on the internal surface of pipes.

Rodding is sending a multi-piece rod into the pipe and removing the precipitates by a reciprocating action of the rod.

Pigging is a similar action, which is done by pushing a traveling scraper object through a pipe. Rodding is generally for smaller bore and shorter pipe sizes while pigging is more for pipelines.

Thus during P&ID development the pipe circuit should be designed roddable or piggable.

Some of the requirements of roddability and piggability are listed in Table 8.8.

Some of these requirements are shown on P&IDs and some of them are only captures as a “note” in the note area of P&IDs.

For solid removal from tanks and other voluminous equipment, as a minimum a “clean-out” door may be needed to be inserted. “Clean-out” doors will be discussed in Chapter 9.

8.9.2 Washing Systems

There are different methods of fluid cleaning.

Flushing can be done for water-soluble scales and fouling. If the scales and fouling are stubborn chemical cleaning using acids, bases or oxidizing agents can be done. If the chemical structure of the fouling is known, and its solvent is economically available, solvent cleaning can be done. Steaming out is common for removing oily fouling from internal of pipes and equipment. It has occurred in the past that right after a steaming out operation inside of a tank, when the steam was still inside the tank, a rapid temperature decrease (for example because of raining) made the steam condense, which created a vacuum that collapsed the tank. There are plenty of companies that

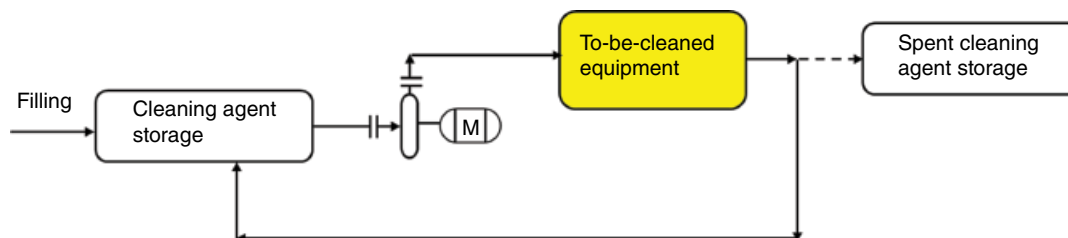


Figure 8.14 BFD of a CIP system.

ask for a design vacuum of a container exposed to steam out to full vacuum to prevent such a disaster.

One example of solvent cleaning is for fired heaters. The internal coil of fired heater can become covered with coke if the to-be-heated stream is a hydrocarbon. From time to time the fired heater can be pulled out of operation and, using recirculating gasoil in the coil, remove the coke created inside the coils.

The fluid cleaning system could be once-through or recirculating. In a once-through system the cleaning fluid is wasted and drained. While in recirculating systems the dirty cleansing stream is sent back, cleaned (e.g. filtered) and is used again.

Cleaning a piece of equipment could be done in any of these forms:

- 1) Fully manual by operators
- 2) The cleansing fluid is hard pipes but the operation is still manual
- 3) The cleansing fluid is hard pipes and the operation is automatic (CIP systems)

The footprint of each of the above forms on P&IDs are:

- 1) Showing only little washing valves on the to-be-cleaned equipment
- 2) Showing washing valves and pipes connected to them
- 3) A CIP system is hard piped to the equipment. A control system connected to switching valves is needed for automatic washing.

CIP systems are the systems that facilitate recirculation of a washing fluid through certain piece of equipment in a plant. Thus, a CIP system includes items like a washing liquid tank, a circulation pump, a spent washing liquid handling system, and finally, a spent washing liquid tank or sump. The washing fluid could be detergent, chemical, solvent, etc. The spent washing fluid could be treated and recycled back as a clean washing fluid if the contaminations in the spent washing fluid are easy to remove in a CIP system. Generally this is the case when the spent washing fluid has only suspended solids. In such cases, the suspended solids can be removed from the spent washing fluid through an in-line filter. However, if the spent washing fluid is so contaminated that cannot be treated easily, it could be dumped into a sump system. CIP systems are used where a piece of equipment or

multiple connected pieces of equipment need to be cleaned frequently and such cleaning may interrupt the operation of a plant.

A block flow diagram (BFD) of a general CIP system is shown in Figure 8.14.

8.9.3 Purging Methods

Purging is an activity used for enclosed spaces to make the atmosphere safe by removing non-innocent gases or vapors.

The purging should be done by neutral gases/vapors like nitrogen, CO₂, steam, or air.

When the process element is voluminous and tall enough, ambient air can be used as the purging gas. This process is named ventilation. In such cases the ambient air is “sucked” through the equipment by a draft. For such cases a sufficient number of manways (minimum of two) and a sufficient size for manways could be needed.

There are some cases where automatic purging is done as part of the safety shutdown system. A common example of this application is in fire heaters. If an emergency shutdown happens in a fired heater, automatic purging helps to empty the heater box of unburnt fuels to make the situation safe.

8.10 Ultimate Destination of Dirty Fluids

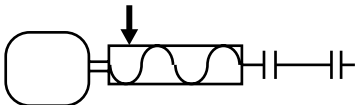
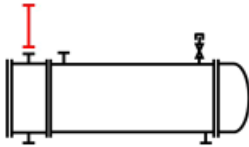
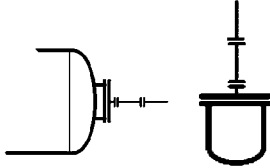
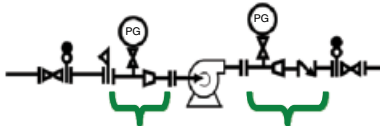
Dirty fluids may come from draining, venting operations or from cleaning operations.

There are two issues to be resolved; one is the final destination of dirty fluids and the second is the way that the dirty fluids get transferred. Both of them are summarized in Table 8.9.

Releasing dirty liquids to the ground was common in older days but with the strict environmental regulations now in place it is no longer acceptable.

These days, by default the drain valves are directed to an “open drain sump” through the surface drainage trench collection network of the plant. The trench network is an open channel network laid down in process areas of the plant, and each releasing point is directed to it by a short piece of pipe.

Table 8.10 Examples of “removal spools.”

Item	Potential location of RS	P&ID example
Progressive cavity pump	Discharge side	
Shell and tube heat exchanger	Tube side	
Vessels/tank	Lines out of flanged head or blinded nozzles	
Centrifugal pump	Suction and discharge side	

to recognize the necessity of RSs in the required locations. Table 8.10 lists a few important RS locations. Sometimes RSs are already present for other reasons on the P&ID (last row of Table 8.10).

8.12 Wrap-up

There is a set of activities that are needed for non-routine conditions of process plants. These are briefly:

- Isolation
- Draining and venting
- Washing and purging.

These activities could be manual or automatic:

- If they are part of shutdown system, they should be automatic.
- If they are part of a rejuvenation process or periodic cleaning they could be automatic (or manual).
- If they are part of an inspection and maintenance operation they are most likely manual.
- If the plan is to remove a piece of equipment from its location (or foundation), “removable spools” need to be provided in suitable locations.

Each of the above intentions need some items to be added on P&IDs.

9

Containers

9.1 Introduction

One important duty of items in process plant is holding and storing process material.

Process materials – for the purpose of holding or storing – can be divided into four groups: flowable solids, non-volatile liquids, volatile liquids, and gases/vapors. Each of these materials may need a specific type of container.

9.2 Selection of Containers

The volume of a stored material is also arbitrarily divided into four classes: low volume, medium volume, high volume, and very high volume.

As can be seen in Table 9.1, for each type of material, in each class of storage, several options are available. Flowable solids can be simply stored in bags if they are of low volume. If the flowable solid is of medium volume, it could be stored in jumbo bag. And if the volume is high it can be stored in a silo. If the solid is not easily flowable and/or it is of huge volume, it can simply be stored in an open pit or on an open pad in the form of a stockpile. Storing in open pits or an open pad can be done only if the weather doesn't destroy the solid.

For non-volatile liquids (like water), if they are of low volume they could be stored in tote tanks. Tote tanks are plastic or metallic tanks with a volume of about 1 m³. Some tote tanks have a volume of 1 m³, some others 1.5 or 1.8 m³. If a liquid of medium volume needs to be stored it could be stored in tanks or vessels. If a non-volatile liquid of high volume needs to be stored the only available option is a tank.

Storing large volume of volatile liquids is done only through using floating roof tanks. A non-volatile liquid can be stored in ponds or reservoirs, as long as it is safe to do so.

For storing volatile liquids, if their volume is low a vessel is available. For medium volume there are two available options of tanks or vessels, and if the volume is high the

only available option is a tank. For volatile liquids, if they are supposed to be stored in tanks, it could be a floating roof tank or a fixed tank with a blanketing gas system.

To store gas, vapor, or highly volatile liquids in low volume a capsule is the available choice. Capsules are a type of vessel that are ended with spherical heads. If gas, vapor, or highly volatile liquid is of medium volume it can be stored in a vessel and if the volume is high a spherical tank is the only option. Gas in very high volumes can be stored only in underground natural reservoirs.

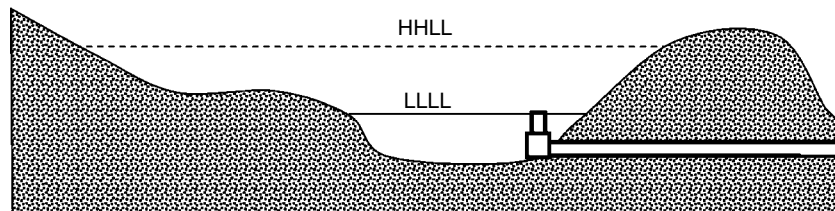
In this chapter we mainly focus on tanks and vessels. The word container can be used for all man-made enclosures in classes of low, medium and high volume. Even though the terms “tank” and “vessel” are sometimes used interchangeably, they are in fact different. Generally, when we are talking about “vessels” it is a container with a high design pressure, while a “tank” refers to containers with a low design pressure, approximately atmospheric pressure. The distinction between low pressure and high pressure is subjective, but usually the division is 15 psig, meaning that if the design pressure of the container is higher than 15 psig it is considered a vessel (or better, a “pressure vessel”), and if the design pressure of a container is below 15 psig, then it is called a tank (or better, an “atmospheric tank”). However, generally the design pressure of tanks is less than 3 psig (approximately 20 kPag or 0.2 barg).

From the above explanation it can be realized that usually containers that have medium or high volume cannot be in the form of vessels, because vessels have a higher designed pressure and the thickness of their body is high. Therefore vessels cannot be built with high volumes economically. If a high volume container is built, it is more attractive from an economic point of view for it to be a tank, which has lower design pressure.

To store very high volumes of non-volatile liquids, ponds, or reservoirs can be used. The depth of ponds is generally limited to 4–5 m (Figure 9.1). A reservoir can be considered as larger and deeper versions of ponds. Reservoirs are common for storing water and it is best to use natural terrain as a reservoir so that minimum change and effort is required.

Table 9.1 Different types of process material storage/holding.

	Low volume	Medium volume	High volume	Very high volume
Solid (flowable)	Bag	Jumbo bag/bin	Silo	Over pads, inside of pits (e.g. landfills)
Non-volatile liquid	Tote tank	Tank (fixed roof)/vessel	Tank (fixed roof)	Ponds or reservoirs
Volatile liquid	Vessel	Tank (floating roof)/vessel	Tank (floating roof)	Not available
Gas/vapor	Capsule (bullet)	Vessel	Spherical tank (vessel)	Underground natural reservoirs

**Figure 9.1** Storm water detention pond.

9.3 Containers Purposes

Containers in process plants are needed for at least three main reasons:

- 1) For (edge) storage
- 2) As intermediate storage
- 3) For unit operations or reactions.

The storage applications of containers are for storing raw material(s) or product(s). This is to make sure there is enough room to stock the products before handing them over to the customer, and also to make sure that raw materials are always available for plant usage. The word “edge” here refers to containers at the beginning or end of the plant. This duty can be termed “long term storage.”

Such containers are generally tanks and have a residence time of a day or days.

The second application of containers in a plant is in the middle of the plant. In this application, they can be used as day containers, feed containers, or surge containers. In intermediate storage the duty is holding or “short term storage.”

Day containers are usually what designers place at the end of each unit to store the material for a “day,” which means anything from 24 hours down to about 8 hours. The purpose of these day containers is to make sure the downstream operations receive the feed even when upstream units are down (out of service) or under maintenance for a short period.

Feed containers are usually placed upstream of “critical” equipment to make sure that a piece of equipment will always receive a continuous flow. For example, we put a feed tank before a furnace or a distillation tower; these units are considered as “critical” because any discontinuity

in feed to these elements may disrupt the operation from a few minutes to a few hours.

Surge containers are placed upstream of critical items to eliminate or dampen small fluctuations in flow to the element. Surge containers are very important in control engineering.

The third application of containers is as a space for “conversion.” Containers are used for different physical or chemical operations. An example of a physical operation is separation, such as distillation or mixing, and examples of a chemical operation are neutralization or reaction. Although vessels are commonly used for any of the three applications of storage, unit operation, and process units, this is not the case for tanks. Because tanks are large containers, they are not easy to control when a physical change or a chemical reaction happens in them. There are few examples of using tanks for the application of unit operations. Tanks can be used for gravity separation of two liquids, like oil from water, and it is not rare to see mixing in a tank through different methods of mixing. Performing reactions inside of the tank is very rare, again because it is difficult to control a huge tank. One very common example of using a tank as a process unit is neutralization tanks. Sometimes a tank is used to receive different acidic and basic strings and neutralization should happen in the tank. Vessels are very commonly used as unit operation and process units. Separation, mixing, and reactions are popular in vessels. Distillation power is a common example of using a vessel as an operational unit.

The residence time for unit operations or process units could be from seconds to hours and more. Biological reactions tend to need higher residence times, up to days.

9.4 Transferring Fluids Between Containers

The duty of containers is storing or holding fluids and also allowing fluids to flow out.

The latter duty is not always an easy one. A similar issue in our daily lives is shown in Figure 9.2.

When transferring fluid from point “A” to point “B,” point “A” should be able to manage a lack of fluid and somehow break the vacuum created, and point “B” should be able to manage the accumulation of fluid and pressurization at point “B.” If either of the points, or both of them, fail to handle the pressure changes, the flow will be stopped. This is one of the most commonly overlooked requirements in developing container systems in P&IDs (Figure 9.3).

From a theoretical viewpoint there shouldn't be such an issue at all. A plant theoretically operates in a steady state and wherever there is flow-in into a container, there should be a flow-out equal to the flow-in. However in practice there are units/plants that are working in batch-wise or semi-continuous modes of operation. Even in fully continuous operation plants there few times that the plant operates in a fully steady state condition. Therefore there is always the chance of liquid accumulation in containers and the creation of low pressure in the source tank and creation of high pressure in the destination tank.

This issue is mainly for liquids, in tanks and not vessels, and specifically for larger tanks.

There should be provision to take care of the atmosphere above the liquid level. Without such provision flow is stopped, or when the liquid level in the container decreases a vacuum will be created in the space in the top of the container. If this vacuum is not broken the container will collapse. In the case that the liquid level in a container increases, the space in the top of the container will be over pressurized. If this overpressure is not released the container will explode.

There are at least four different ways to deal with this issue. They are explained below and are shown in Figure 9.4.

Solution 1: do nothing. This solution can be used when the amount of vacuum or overpressure is very slight and at the same time the liquid is near its boiling point. This means the liquid can easily be converted to vapor and vapor can be easily converted to liquid. In this solution the slight vacuum created will be compensated for by additional liquid evaporation and the slight overpressure will be mitigated by a small conversion of vapor to liquid. This solution is not very reliable and is rare.

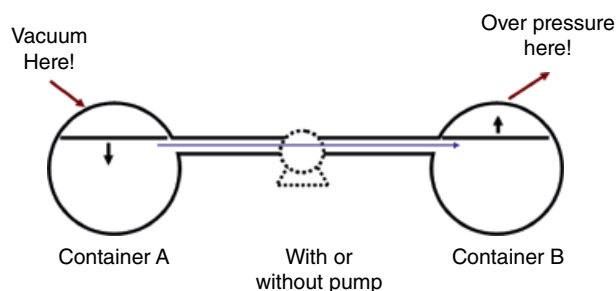


Figure 9.3 Fluid transfer between containers.

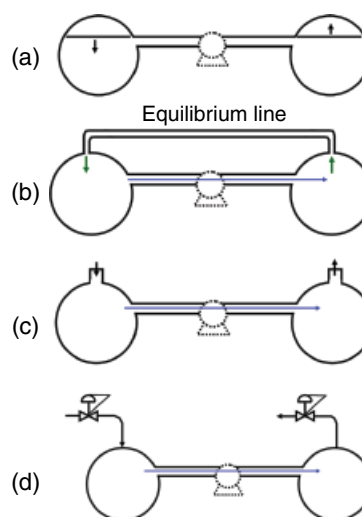


Figure 9.4 Different ways of facilitating flow.

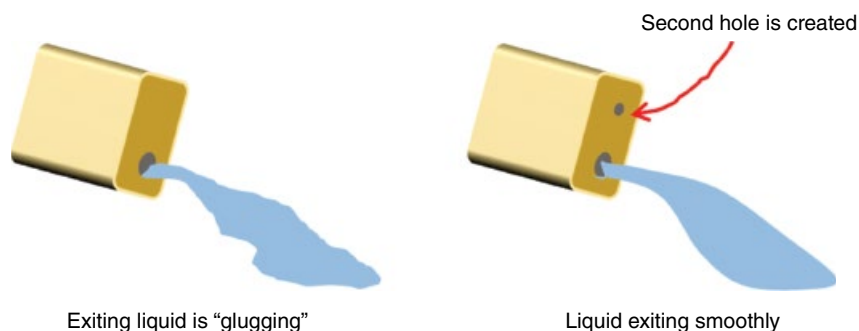


Figure 9.2 Problem of pouring liquid out of a can.

Solution 2: vent to atmosphere. When a liquid is moved from a source container to a destination container there will be no issues if both of them are equipped with free vents. In the source container no vacuum will be created because the ambient air will be sucked into the container and there will be no creation of overpressure in the destination tank because the compressed vapor goes out of the container to the atmosphere through the vent. This solution is very common if a vent to atmosphere is suitable and the presence of air in the vapor space of the container is acceptable. For example this is good solution for water tanks where the “emission” of water vapor won’t violate any environmental regulations and air can exist in the vapor space of the water tank. However, it is probably not an acceptable solution for a liquid ammonia tank.

Solution 3: pressure balance pipe. In this solution a piece of pipe connects the vapor space of container “A” (source container) to the vapor space of container “B” (destination container). Using this trick the liquid flows from container “A” to container “B,” the liquid level in container “B” will be increased, and the vapor is pushed out of container “B” and through the balance pipe toward container “A.” Container “A” needs an entrance for vapor otherwise a vacuum would be created in container “A.” A pressure balance line or equilibrium line is only practical if the two containers are close to each other.

Solution 4: compensating gas. In this solution the vapor space of both containers, source container and destination container, will be connected to a source of gas. The flow of these sources of gas are regulated by two regulators or control valves. These pressure control valves or pressure regulators allow compensating gas to go through the source tank to prevent the creation of a vacuum. The pressure regulator on the vapor space of the destination container will open to release the vapors out of the destination container to prevent the creation of overpressure. This compensating gas for the source container is basically the same as blanketing gas. The compensating gas in the destination tank is basically the vapor to the vapor recovery unit (VRU) network. Blanketing systems will be discussed further in Section 9.12.

9.5 Container Positions

The container’s position in relation to the ground level can be different (Figure 9.5). It can be classified as “underground,” “in-ground,” “above ground,” and “elevated.”

A container installed and buried in the soil is named an underground container. Underground containers have the issue of limited accessibility for monitoring and

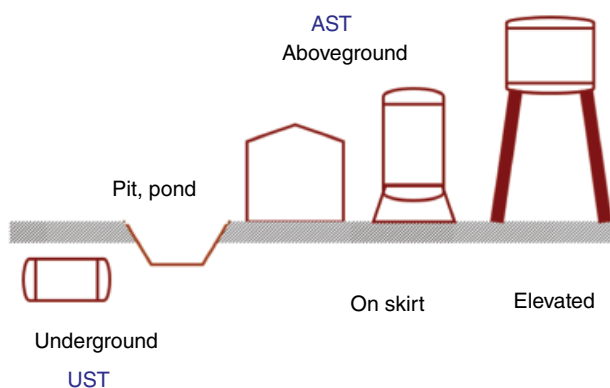


Figure 9.5 Position of containers.

inspection. Regulatory bodies generally don’t like underground containers because of the chance of leakage of liquid content to soil and eventually to underground water. Underground containers could be used when the incoming flows don’t have enough pressure (like sumps) or where there is a chance of being hit by moving objects around them. One famous example of an underground container is gasoline containers in gas stations. Gasoline containers in gas stations are almost always underground because a gas station is an area in which drivers (not necessarily skillful drivers) are maneuvering. For this reason, the gasoline container is always underground in gas stations. When an underground container is necessary there are ways to address the regulatory bodies’ concerns. One solution is using double wall containers. The other solution is installing the container inside of a pit. In this case, the container is underground but not buried. Underground containers should be signposted by placing barricades on the ground. This is generally so that vehicles do not place heavy loads on the ground that has the buried container.

The other containers are installed in such a way that the top of their edge is at the same level as the ground and these are named in-ground containers. These containers can be named as a pit if they are very small, basin if they are of a decent size, or a pond/reservoir if they are huge.

Containers can be installed on the ground surface and are named above ground containers. Above ground containers could be installed directly on foundations, or on a skirt, or on legs. The best container installation is the on-ground container, which is generally named an above ground container. They are easy to install, easy to monitor, and easy to inspect. The above ground vessels, either vertical or horizontal, are installed on legs or a skirt. Vertical tanks or silos can be installed directly on foundations provided by civil engineers or on legs/a skirt. As a rule of thumb, if the diameter of the tank is larger than 5 m it should be installed directly on the foundations

Table 9.2 Advantages and disadvantages of above ground and underground containers.

	Underground	Aboveground
Pros	<ul style="list-style-type: none"> • No chance of vehicle impact • As destination, no need for pumping • Less chance of freezing if buried below the frost line 	<ul style="list-style-type: none"> • Less expensive • Easy inspection
Cons	<ul style="list-style-type: none"> • Chance of leakage and loss of containment • Higher capital cost • Difficult to inspect 	<ul style="list-style-type: none"> • Chance of vehicular impact • They reduce accessibility to areas
Examples	<ul style="list-style-type: none"> • Flammable liquids • Drain sumps 	<ul style="list-style-type: none"> • By default choice

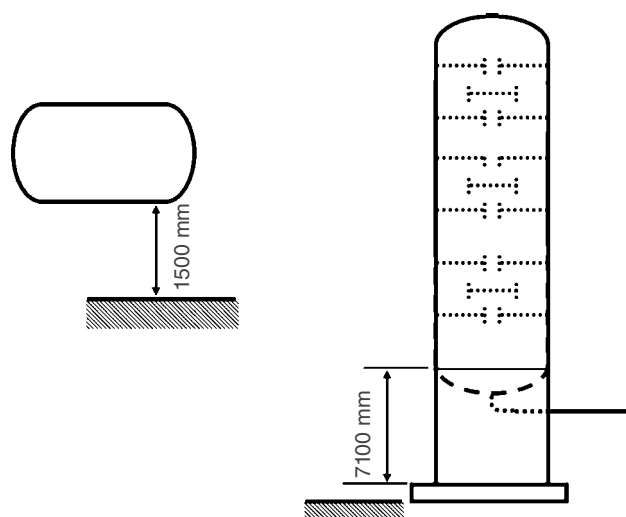
and if the diameter is less than 3 m it should be installed on skirt/legs. However, this is only a rule of thumb; there are some cases that a large diameter tank is installed on legs and the client has agreed to pay extra money for that.

The last type of container installation is an elevated container. An elevated container is a type of container that is installed on long legs. Because of the different issues related to elevated containers they are not very common in process industries. It is difficult to monitor and inspect an elevated tank and there is always that chance of it being hit by flying objects. Elevated tanks are mainly used only for storing raw water or potable water to provide specific high pressure at the outlets of the container.

Arguably the two very popular positions of containers are above ground and underground.

Table 9.2 shows some advantages and disadvantages of underground versus above ground tanks.

There are some cases where the process engineer dictates a specific elevation for a container. In such cases the elevation should be stated on the P&ID (Figure 9.6).

**Figure 9.6** Elevation of containers.

9.6 Container Shapes

Containers are generally in one of three shapes: cylindrical, cubical, or spherical (Figure 9.7).

Tanks could have the shape of cylindrical or cubical.

Vessels generally have the shape of cylindrical or spherical.

The most common type of container shape is cylindrical. Cylindrical vessels can be installed in horizontal or vertical positions. Cylindrical tanks are obviously installed in the vertical position.

Another shape of container is spherical. Spherical tanks are basically pressure vessels, and are used only for high volume storage of gases or vapors or high volatile liquids. The reason for the less common usage of spherical tanks is they are expensive. The design and fabrication of spherical tanks are both expensive.

The last shape of container is cubical. As a general rule, metallic containers are preferably built in cylindrical shapes and concrete containers in cubical or cylindrical shapes. Using metallic containers in the forms of underground or in-ground containers is generally avoided, to avoid the contact of metal with soil. This is to prevent corrosion.

Using concrete containers in the form of underground or in-ground containers is excellent. On the one hand there are fewer issues of corrosion and on the other the surrounding soil supports the concrete slabs and a thinner slab may be required.

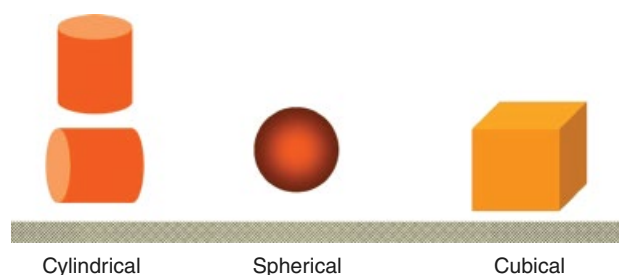
**Figure 9.7** Different shapes of containers.

Table 9.3 Features of different container shapes.

Cubical	Cylindrical	Spherical
Easy to build specially concrete containers	Easy to build	Very difficult to build and install
Can be put side by side to save the cost of building extra walls	May need a larger footprint than side by side cubical containers	May need a larger footprint than side by side cubical containers
Needs thickest wall thickness in similar conditions	A medium size of body thickness is necessary	Needs the thinnest wall thickness in similar conditions
In the role of units, the best matches for plug regime units	Can be used for completely mixed units	In the role of units, the best matches for completely mixed units
Very common for concrete basins	For metallic, polymeric containers	Mainly for metallic containers

The different features of container shapes are listed in Table 9.3.

9.6.1 Closing Parts of Containers

Except for spherical tanks, where one word represents the whole shape of the walls around the stored process material, other container shapes should be clarified by at least two words: the shape of the main portion and the “closing” portions. For example what is known as “cylindrical vessel” introduces only the main portion of the container. The closing parts of cylindrical vessels are “heads.”

When we are talking about cylindrical containers we need to talk about both sides of the cylinder. In a vessel each side of a cylinder has a “head.” For tanks, at the top side of the cylinder is called the “tank roof” and the bottom side is called the “tank floor.”

Tank roofs and floors will be discussed in Sections 9.15 and 9.17.

If it is decided that a container is to be open topped, it doesn’t need a closing part. For other containers closing parts are needed.

9.6.2 Open Top or Fully Enclosed Containers

One decision regarding vertical containers (tanks or vessels) is that the top of them should be closed or open.

The decision of an open top container definitely provides some cost saving but it has some disadvantages too. Open top containers allow contact between stored material and the environment. Such contact could be harmful for each or both of them.

However open top containers allow the operators to monitor and inspect the system very easily and frequently. This is one reason that the majority of containers in water treatment plants are open top. This excludes the last container, the treated potable water container, which obviously should be fully enclosed.

The different features of open top and fully enclosed containers are outlined in Table 9.4.

Open-top containers are not very common in the chemical industry.

The decision to fully enclosed large containers like ponds and reservoirs needs detailed study. Sometimes when a large pond needs to be fully enclosed (roofed), it loses its attractiveness. Such reasons are generally code requirements.

9.7 Container Identifiers

Based on the concepts stated in Chapter 4, the identifiers of containers in P&IDs are symbol, tag, and call-out.







9.7.1 Container Symbol

These container symbols can be seen in Table 9.5.

Table 9.4 Open top versus fully enclosed containers.

	Open top	Fully enclosed
● Reasons	<ul style="list-style-type: none"> ● To save money ● When frequent monitoring is needed 	<ul style="list-style-type: none"> ● When the contained fluid is at high pressure, or emits harmful vapors ● When the stored fluid can be contaminated when in contact with the environment
● Applicability	<ul style="list-style-type: none"> ● Applicable to in-ground, above ground, or elevated containers ● Not applicable for underground containers ● Not common for elevated containers 	<ul style="list-style-type: none"> ● Applicable to in-ground, above ground, or elevated containers ● A definite need for underground containers

Table 9.5 P&ID symbols of different containers.

Type	Symbol		
Tanks	Open top	Fixed roof	Floating roof
			
Vessels	Cylindrical horizontal	Cylindrical vertical	Spherical
			

9.7.2 Container Tags

The necessity of putting a container tag on the body of P&ID is mentioned in the project documents. If container tags need to be shown on the body of P&IDs, they are generally placed inside of containers unless there is not enough room or it causes confusion.

As was mentioned in Chapter 4, the tank tag could be “T-3420” or “TK-3420.” For a vessel the tag could be “V-232.”

9.7.3 Container Call-outs

Container call-outs can be different depending on the type of container, tank or vessel, and the usage of containers, conversion units or holding.

9.7.3.1 Tank Call-outs

When assigning a call-out for a tank, at first it should follow the company’s guidelines. However, there are some general guidelines that are mentioned here.

A typical call-out for a tank is shown in Figure 9.8.

Here we explain the items in Figure 9.8.

- The first line is the tag number of the tank.
- The second line is the name of the tank. The name of tank is generally decided based on the tank content.

600-TK-101
Boiler feed water tank
Size: 8,850 mm ID x 9,000 mm H
Capacity: 550m³
Design: 3.45 KPag @ 80°C
0.25 KPag VAC. @ 80°C
Material: CS
MDMT: -29°C
Insulation: 38H
Trim: TT-419-DAB

Figure 9.8 Call-out of a tank.

- The third to fifth lines are the main characteristics of the tank; dimensions and capacity is stated. What is referred to as capacity? In regards to containers it is in essence the total volume of the container. “Total” here means all the container space, even including the space that is not generally filled with liquid. It is important to recognize the difference between the values of capacity in P&ID versus the capacity that is mentioned in the process flow diagram (PFD). In PFDs capacity refers to the portion of the container capacity that is usable during “normal” operation, while in P&ID capacity means the total physical capacity. Containers can also be specified by their dimensions.
- The next lines are the design pressure and temperature. The other important items in a container call-out are the designed pressure and the designed temperature. Whenever we talk about designed pressure for an enclosure, it could be two types of pressure, internal pressure and external pressure. Therefore, for containers we need to specify the designed pressure and the designed vacuum. Design pressures and temperatures will be discussed more in Chapter 17.
- Material of construction could be the next item on the call-out. In this line the general names of the material are mentioned without going through the detail of the material.
- Minimum metal design temperature (MDMT) could be another item in a container call-out, which is the acronym for “minimum designed metal temperature.” The MDMT is basically the minimum temperature that the wall of a container may “see” during its lifespan. To begin with the concept of MDMT was used for metallic containers only, but nowadays it is also applicable to non-metallic containers like FRP containers.

- Then, if the tank has any internal and/or external attached layer, it could be stated here. Examples are insulation (on the outside of the tank), lining (inside of the tank), cladding, etc. If there is anything to roughly specify them, it could be mentioned. For example in Figure 9.5, it is specified that there is insulation for the purpose of heat conservation and its thickness in 38 mm (or 1.5"). Insulation will be discussed more in Chapter 17.
- Trim could be another item in the tank call-out. First of all we need to answer the question of what trim is. "Trim" is basically all the little appendices attached to a piece of equipment that is in the form of short pipes. For example trim for tanks is all the nozzles connected to the body or roof. The trim's material is not necessarily the material of the rest of tank. The material for trim should somehow be identified. "Trim code" or "trim" is basically is a code string that is assigned to the trim of a tank to specify their material. The trim code is very similar to pipe tags.

9.7.3.2 Vessel Call-outs

Vessel call-outs are not very different from those of tanks. A typical call-out for a tank is shown in Figure 9.9.

The items in vessel call-outs are very similar to those in tanks. Here we talk about the items that are different.

The size of vessel is stated as one of the items in call-outs. In Figure 9.9 the dimensions of a cylindrical vessel are given. There is no complexity in the diameter of vessel, which is 4000 mm internal diameter. The length (or height) of vessel could be stated in at least in two ways. This is because the cylindrical vessels are completed with two closing heads and the heads are not flat sheets.

If stating the vessel's length only the length of cylindrical portion is considered and is shown as "S/S." "S/S" represents seam-to-seam, which refers to the seam lines at both ends of cylindrical portion.

Each closing head has a straight portion. Therefore if the length of vessel is considered to be the full length of the straight portion, the length is shown as "T/T." "T/T" represents tangent-to-tangent (Figure 9.10).

As a rule of thumb, the straight portion of the heads is about 1.5"–2" irrelevant of the head diameter. Therefore

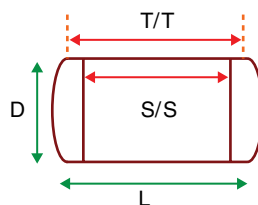


Figure 9.10 The two ways of stating vessel length.

the vessel length in T/T is about 3"–4" longer than the vessel length in S/S.

Another item in the vessel call-out is the "design pressure at design temperature." In Figure 9.9 the design pressure (internal pressure) is given as 970 KPag at the design temperature of 70 °C and "FV" or "full vacuum" (external pressure) at the design temperature of 120 °C. "FV" is nothing other than a design external pressure of –101.3 KPag. "FV" is a very common design external pressure for small and mid-size vessels. The reason the "FV" may not be used for large size vessels is because of the huge extra cost. A large vessel with an "FV" design measure could be very expensive and not affordable. In Chapter 17 there is more discussion on design pressure and design temperature.

Although the design external pressure could be selected to be any value from 0 to –101 KPag, but in practice the designers do not bother to calculate the exact required design external pressure and rather may select "FV" (–101 KPag) or "HV" (–50 KPag), "half vacuum."

Design pressure and design vacuum of containers are not closely related to each other. For example in water bottles. It needs very high pressure to exploding them but you can easily collapsing them!

Material of construction for a single container or a piece of equipment is not necessarily unique. Thus in some cases more than one trim is stated in the call-out (Figure 9.11).

Such thing may happen for containers in the role of unit operation or a process unit.

300 V-4125
Reflux drum
 Size: 4,000 ID mm x 5,000 mm T/T
 Design: 970 KPag @ 70°C/FV @ 120°C
 MDMT: –29°C
 Insulation: H/50 mm
 Trim: 565NG2167 -BAS
 Lining: epoxy

Figure 9.9 Call-out of a vessel.

F-122
Distillation tower
 Size: 8' - 0" DIA. x 38' - 0" H (T/T)
 DP @ DT: 100PSIG @ 80°F
 Material: CS
 Trim: AC-112 (above tray 10)
 Trim: BC-115 (below tray 10)

Figure 9.11 Multiple trims in a distillation tower.

F-3100-C
Sand filter
Design: 1000 KPag/FV/120°C
Design capacity: 400 m³/hr
Size: 3600mm I.D. × 2889mm H (S/S)

Figure 9.12 Call-out of a sand filter.

9.7.3.3 Tag of Container in Duty of Conversion

The tag of a container in the duty of physical or chemical conversions (unit operation or process unit) can be tagged as a minimum based on their dimensions, similar to simple containers.

However, some companies add more mechanical detail to them in their call-out. “Design capacity” (in flowrate units) could be one of them (Figure 9.12).

9.8 Levels in Non-flooded Liquid Containers

In Chapter 5, we talked about different levels for each process parameter. We can define such level system for liquid levels in non-flooded liquid containers. The liquid level is the only process parameter for which levels should be mentioned on the P&ID for each non-flooded container.

These liquid levels are traditionally normal level, high level, high-high level, low level, low-low level. The liquid levels are defined by the process engineer during the design phase.

It is important to know that we do not always have these five levels for liquids. For example, in raw material tanks generally there is no normal liquid level (NLL). Raw material tanks can be filled out in tankers or ships up to one high level to make them full, and after usage the tank content drops to the level where eventually requests for filling out the tank should be placed. Therefore, here we have high liquid level (HLL) and low liquid level (LLL), but there is no normal liquid level (NLL) per se in raw material tanks.

One other example for non-traditional notification of levels in tanks is liquid two-phase fluid in tanks. When there is a two-phase liquid fluid like oil and water in a tank the location of the interface could also be important. In such cases we may have one additional level as NIL (Normal Interface level).

The third example of having non-traditional level notifications on tanks is containers that are connected to set of pumps in a lag-lead arrangement. In such arrangements two or more pumps exist in the outlet of a container but they do not necessarily operate simultaneously. The second pump (lag pump) may start to operate when the level in the container gets to a high level, which means the first operating pump (lead pump) cannot keep

up to decrease the liquid level. In that case the second pump starts to operate to help the first operating pump. In this example we may have some additional liquid levels on the tank symbol in P&ID and they could be “lag pump on level” and “lag pump off level.”

9.9 Container Nozzles

For containers to connect to pipes they need nozzles. Nozzles are basically the “hands” of containers that allow containers to be connected to a piping system. The nozzles of containers are provided not only for the normal operation of the container but also for non-normal operation, maintenance, and inspection.

Nozzles are shown on containers in Figure 9.13.

As it can be seen, nozzles can be shown as plan view or section view. In some P&IDs, a nozzle number could be assigned to each nozzle. This numbering facilitates connecting the right pipes to the right nozzles during the construction of a plant. The nozzle numbers that are used in the P&ID are in accordance with the nozzle number in the container datasheet. They could be shown as a hexagonal symbol beside each nozzle in the container. It is very common to number nozzles as N1, N2, N3, etc.; however, there are some special nozzles that can be specified by a more specific numbering system. For example “MW” means “manway,” which is a nozzle and container for entering the container when it is empty during inspection or maintenance. “TH” is another common acronym on some nozzles, which represents “thief hatch.”

There are different questions that should be answered regarding the containers nozzles. They are: how many nozzles should be put on a tank? Should a nozzle be placed on the tank roof or tank shell? What should be the size of each container’s nozzle? What are the different duties of the nozzles? These questions are answered one by one in Section 9.9.1.

9.9.1 Nozzle Duties

There are different ways that nozzles can be classified.

Nozzles can be classified as process nozzles and instrument nozzles. Process nozzles are for the purpose

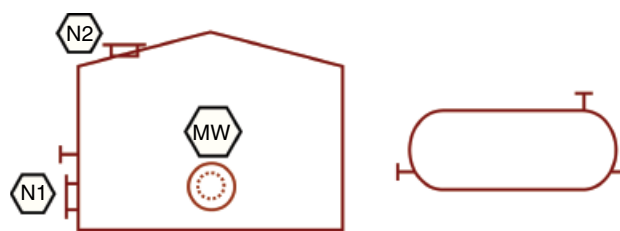


Figure 9.13 Showing container nozzles on P&IDs.

of process and instrument nozzles are related to instruments. They are nozzles that are installed on a container to attach sensors to a container. The sensors could be level sensors or temperature sensors.

Nozzles can also be classified as operating nozzles and spare nozzles. Operating nozzles are the ones currently in use. Spare nozzles are implemented for future potential use.

Table 9.6 summarizes the name and purpose of several container nozzles.

The size and the number of spare nozzles could be specified by the client. Otherwise the process engineer can use his judgment. For process spare nozzles he/she

can use a rough number of 10–20% of operating process nozzles, with sizes in the mid-range of existing process operating sizes. For instrument spare nozzles several 2" and 3" nozzles can be provided.

9.9.2 Nozzle Locations

It is important to know that the location of nozzles on containers in the P&ID never ever represent their real location on containers. The real location of nozzles on each container can be found in the container datasheet. The nozzle symbols are drawn on the container symbol just wherever there is enough room around the container.

Table 9.6 Duties of nozzles.

Nozzle name	Explanation
Process fluid nozzles	Process nozzles are nozzles that process fluid goes through for entering the container fluid or exiting the container fluid
Instrument nozzles	Instrument nozzles are the nozzles for installation of different sensors to the containers
Manway	Manways are not only for quality control or maintenance personnel to enter. Manways are used for operator entrance and also entering or exiting container internals. They are generally used in large containers, ones with a diameter more than one meter
Overflow nozzle	Overflow nozzles provide a release route when the liquid level in a tank keeps increasing and is going to fill the container
Thief hatch	Thief hatches used to be called dip hatches. They are nozzles with easy-to-open lids and are installed on the top of liquid containers (for example on the tank roof) for the purpose of taking samples from the container liquid. This sampling is done by sending down (dipping) a small jar to the liquid. In older times, these nozzles were used by thieves to steal oil products from tanks. This is the reason their name gradually changed from dip hatch to thief hatch
Pressure protection nozzle	This is a nozzle (or nozzles) that releases fluid out of the container to protect it against high pressure. Basically, this nozzle is for the installation of a PSV
Vacuum protection nozzle	This is the nozzle (or nozzles) that provide contort release of fluid out of the container to protect it against high vacuum. Basically, this nozzle is for the installation of a VSV
Free vent nozzle	Free vent nozzles can be necessary for non-flooded containers. They are installed mainly on atmospheric tanks. These are used to provide an opening for tank "breathing." "Breathing" will be discussed in Section 9.11
Clean-out doors	Clean-out doors are generally larger nozzles of mainly rectangular shape that are placed on large containers (large tanks) to provide capability of sending machine to the tank. For a plenty of cases, these machines are for sludge removal from the bottom of the tank.
Drain nozzles	Drain nozzles are the nozzles used to drain liquid from a container for the purpose of taking out the container from normal operation.
Hand holes	Hand holes are nozzles on smaller containers that are too small for an operator to be able to enter into. Hand holes are used only for inserting a hand to check the inside of container. They are used for smaller containers instead of manways.
Drain and vent size	For draining and venting the container
Sting nozzle	For cleaning during the operation
Steam-out nozzle	For off-line cleaning
Purge nozzle	For off-line cleaning and making safe the internal atmosphere
View or inspection port	For monitoring by the operator
Media adding/removal	For adding or removing media such as sand or resin beads during rejuvenation
Cathodic protection	For corrosion protection

A big mistake is sometimes made by piping modelers in that the containers are modeled based on their P&ID symbol rather than the container datasheet. Possibly the only exception to this general rule is for nozzles on a tank shell or on a tank roof. If a nozzle should be on a tank roof it should be drawn on the tank symbol in the P&ID and if a nozzle should be on a tank shell it should be drawn on the tank shell in the P&ID. Here lack of room to show a nozzle on a shell is not a good excuse to put a shell nozzle on the roof of the tank.

Tanks have roofs and shells, and their nozzles could be on the roof or on the shell.

We like to install the nozzles in a location that is easy for monitoring and inspection. It is easy to recognize that nozzles on the shell are more accessible than nozzles on the roof. Plant operators don't like roof nozzles because they need to take steps up to be able to inspect them. Therefore the general rule is to put all the nozzles as much as possible on the shell of a tank. However, there are some nozzles that cannot be installed on the tank shell. They are for example vent nozzles, PSV nozzles, VSV nozzles, blanket gas and VRU nozzles (discussed at the end of this chapter) and Nozzles for none-contact type level sensors. The none-contact type level sensors will be discussed in Chapter 13.

If there are nozzles to be placed on the shell, it is best to put them on the lower part of shell, which is more accessible. If there are nozzles to be placed on the roof, generally they are grouped into one area to be accessible by installing only one catwalk over the roof. However, these requirements cannot be met in all cases. For example overflow nozzles are always located at the top of the tank's shell.

Contact type level instruments should be placed in the middle of the shell where the liquid level is.

Table 9.7 summarizes the features of shell and roof nozzles.

Table 9.8 Nozzle locations for tanks.

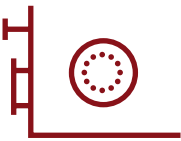
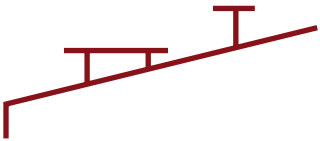
Nozzle Name	Location
Process fluid nozzles	Preferably below LLLL
Instrument nozzles	Lower than LLLL, except level sensor
Manway	Different locations
Overflow nozzle	On the top of shell
Thief hatch	On top or roof
Pressure protection nozzle	On top or roof
Vacuum protection nozzle	On top or roof
Free vent nozzle	On top or roof
Clean-out doors	On the bottom of shell
Drain nozzles	On the bottom of shell
Truck-out nozzles	On the bottom of shell
Hand hole	Within the reach of operator
Drain and vent size	On the bottom of shell
Sting nozzle	On the bottom of shell
Steam-out nozzle	On the shell
Purge nozzle	On top or roof
View or inspection port	Within the reach of operator
Media adding/removal	On the bottom of shell
Cathodic protection	On the bottom of shell

The general location of some nozzles is listed in Table 9.8.

9.9.3 Nozzle Elevation Versus Liquid Levels

In non-flooded liquid containers it is important to note the relative elevation difference between nozzle elevation versus liquid level in the container.

Table 9.7 Nozzle locations for tanks.

	Nozzle on shell	Nozzle on roof
Schematic		
Features	<ul style="list-style-type: none"> • More accessible • Chance of liquid leakage • More prone to clogging 	<ul style="list-style-type: none"> • Less accessible • Leak-proof against liquid • Could be more prone to corrosion
Examples	By default choice	Others

As a general rule, we try to put the nozzles below the LLLL (low-Low liquid level) as much as possible. However, there are some important exceptions.

There are some nozzles that should be installed at specific elevations to function properly.

For example contact type elevation sensors should be installed at specific elevations to be able to sense liquid levels in the wide range of elevations. A hydrostatic type level sensor – which has two nozzles – can be installed in such a way that the lower nozzle is somewhere below the LLLL and the top nozzle is above the HHLL.

Another example is a conventional overflow nozzle. A conventional overflow nozzle doesn't have any internals. This nozzle should be installed somewhere above the HHLL, possibly 30 cm above the HHLL.

The other nozzles tend to be installed below the LLLL to make sure they always “see” the liquid. The bulk of liquid below the LLLL down to the container bottom (or floor) is called “dead liquid” as it is always there.

The other reason that we don't like nozzles on the top of tanks is the exerted stress on the tank. The sheet thickness on the top of tanks is thinner than the bottom, so it may not be able to tolerate the high weight of nozzles.

The inlet process nozzles are specifically installed on the lower portion of the tanks. If they are on the higher portion of tanks, the incoming flow creates vibration in the tank and also non-necessary agitation in the tank liquid, which may accelerate corrosion and/or generate static electricity (Figure 9.14).

However, there are some cases that process reasons may dictate installing the inlet process nozzles on the higher portion of tanks.

One example is in some “aggressive” raw material tanks that are filled by haulers. In such cases if the filling nozzle is placed on the bottom portion of the storage tank, the filling pipe will remain filled with stagnant aggressive liquid all the time. This may cause problems for the system and/or the operator. If the liquid is a slurry it may plug the filling pipe. If the liquid is hydrochloric acid, it will splash when the operator tries to connect the quick connection. Because of all these reasons it is preferred to leave the pipe empty after each offloading. This could be done by placing the inlet

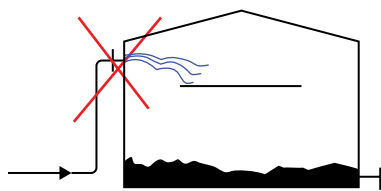


Figure 9.14 Problems with the elevated inlet in non-flooded containers.

nozzle on the top portion of the tank. After each offloading the piping could be drained to make sure no liquid remains in the piping.

The offloading inlet nozzle is equipped with an internal down comer extension to prevent the agitation in the tank liquid (Figure 9.15).

The other example is when a pump should pump to a tall tank. In such cases if the inlet nozzle is placed at the bottom of the tank, the pump should be able to pump liquid to the tank when the liquid level is at its highest elevation or at its lowest elevation. Sometimes it is difficult to find a pump to operate on its curve (head–flowrate curve) in both cases while it is in a good efficiency area. In such cases the inlet nozzle can be placed on the top portion of the tank. This problem is not necessarily just for tall tanks. A tank that looks “conventional” may also experience this problem if the difference between its liquid levels is large (Figure 9.16).

The other example is when dealing with two phase flow of gas–liquid. A two phase flow inlet nozzle at the bottom of a tank causes sparging of gas into the liquid bulk of the tank. To prevent this problem the nozzle could be placed on the top portion of the tank to allow the gas–liquid flow to disengage to gas that goes up and liquid that drops to the liquid bulk (Figure 9.17).

This arrangement also helps when we are dealing with a liquid with dissolved gas. The dissolved gas will be partially stripped off when flow splashes above the liquid level.

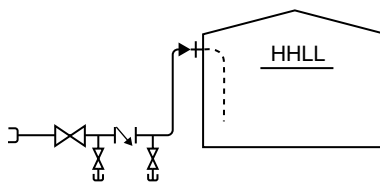


Figure 9.15 Required elevated outlet nozzle in intermittent, operator involved applications.

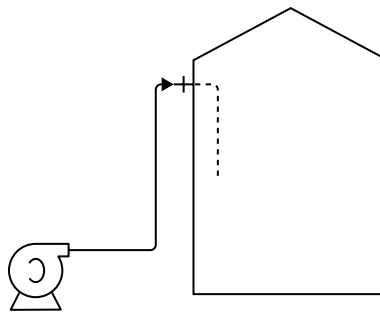


Figure 9.16 Required elevated inlet nozzle when the pump cannot handle both low and high levels.

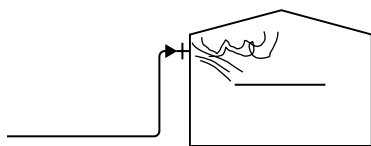


Figure 9.17 Preference of elevated inlet nozzle for two phase liquid-gas flow pipes.

9.9.4 The Size, Number, and Rating of Nozzles

Nozzles sizes and pressure ratings are important, too.

Nozzle sizes can be seen on P&IDs if they are not connected to a pipe or are different from the connected pipe size.

Nozzle numbers are decided based on the required functionality.

Nozzle pressure ratings are only shown on the P&IDs IF the pressure rating is different from that of the pipe that is the connecting pipe to the nozzle.

The pressure rating of nozzles should match the design pressure of the container they are attached to.

However we generally don't choose a 150# pressure rating for small nozzles (say less than 3") even though they are connected to atmospheric tanks. To provide enough integrity in small nozzles the minimum pressure rating for them is chosen to be 300#.

Typical sizes of different nozzles are shown in Table 9.9.

9.9.5 Merging Nozzles

The number of nozzles on containers should be kept to a minimum. One reason is nozzles are expensive parts and the other reason is that nozzles are heavy, and installation of more than needed nozzles on containers may force us to use thicker and more expensive shells for container fabrication. There are different ways to decrease the number of nozzles on containers. For example, if there are several incoming fluids to a container, instead of a locating one nozzle for each stream we may merge the streams together and then direct the mixed flow to the container through one nozzle (Figure 9.18).

Table 9.9 Size of nozzles.

Nozzle name	Size	Number
Process nozzles	Minimum 2" (to avoid plugging)	Per connecting pipes' number
Instrument nozzles	2"–3"	Per required instruments
Manway	24"–30"–36"	Per client request or one manway per every 10 m of diameter of container for shell manways, and for roof manways one manway per every 15 m of diameter
Overflow nozzle	Needs sizing As a rule of thumb: maximum inflow size	One
Thief hatch	The same as manway size	One (or occasionally two)
Pressure protection nozzle	Needs sizing	Needs sizing
Vacuum protection nozzle	Needs sizing	Needs sizing
Free vent nozzle	Needs sizing	Generally one
Clean-out doors	8" × 16", 24" × 24", 34" × 48", 48" × 48" (Limited application)	One
Drain nozzles	Up to 2"	Multiple
Truck-out nozzles	3"–4"	Multiple
Hand hole	8"	Number to provide full reachability
Drain and vent size	Discussed in Chapter 8	Multiple
Sting nozzle	2"–4"	Multiple
Steam-out nozzle	1½" or the same as vent size	Multiple
Purge nozzle	The same as vent size or one size larger	One or two
View or inspection port	6"–8"	One
Media adding/removal	4"	One
Cathodic protection	4"–6"	Needs sizing

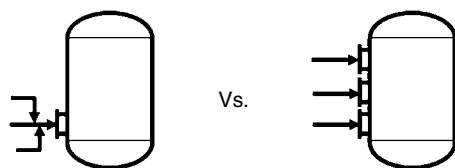


Figure 9.18 Decision on the number of process nozzles.

However, this solution has some shortcomings. In this solution there is always the chance of reverse flow in one or more tying pipes. If there is a chance of intermittent flow in one of the tying streams, or the chance of dropping pressure in any of these streams, a check valve should be installed on the intermittent or low pressure stream.

The other disadvantages of having more than enough nozzles are that each nozzle potentially is a source of leakage or source of corrosion. However, by shifting a nozzle from shell site to roof site the chance of liquid leakage is eliminated, but this is not always doable.

The other solution to minimizing the number of nozzles on containers is using shared nozzles rather than the dedicated nozzles. For example, some companies use one nozzle on top of the tank shell for the dual purpose of venting and overflowing. This is important to consider that two nozzles can be merged to one shared nozzle if the functionality of each of them is not overlapping. It means the need for each nozzle should be only in one phase of operation. In the previous example where we merged overflow nozzle and vent nozzle it was doable because the vent nozzle needed to be functional during the normal operation of a tank while the overflow nozzle only needed to be functional during upset operation. Therefore, these two nozzles could be merged to a single shared nozzle because they are not operating in the same phase. The other examples for using shared nozzles are using one nozzle for flushing and draining a container. The other example is using manway nozzles for the purpose of creation of draft to dry up the tank internally for maintenance purposes. One example is the nozzles that have the capability of being a PSV and a manway at the same time, or nozzles that are a PSV and “thief hatch” at the same time.

9.9.6 Nozzle Internal Assemblies

Nozzles on containers can be installed without any internal connected assembly and this is in the majority of cases. However, there are some cases that a nozzle should be connected to an assembly from the other side and inside of the container. The most common types of nozzle internal assemblies are vortex breakers, down comer, risers, and extensions, which are shown in Figure 9.19.

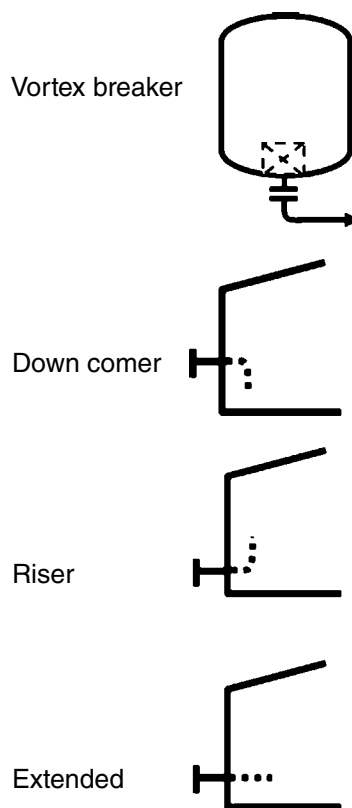


Figure 9.19 Nozzles internals.

Vortex breaker: Vortex breakers are the type of nozzle that is sometimes installed on some outlet nozzles of liquid containers. The main purpose of the vortex breaker is to prevent the creation of a vortex in the outlet of a container. You may have seen the creation of vortex when you try to empty your bathtub. The vortex that we may see in our bathtub is not very harmful but in an industrial context it should be avoided. The reason we do not want a vortex in the outlet nozzle of some liquid containers is that it causes some erosion inside the container and also the atmosphere inside the tank may be entrapped in the outgoing liquid, which may not be a good thing. Prevention of vortices is more important when a liquid container sends liquid to the suction of a pump. If prevention of vortices fails the liquid going in to the pump suction may have some gas bubbles, which is detrimental for the pump operation. Therefore, it is important to know that vortex breakers may be needed only if it is a liquid container and on an outgoing nozzle. One important thing is that not all outgoing liquid nozzles need vortex breakers. There are some formulas that show whether we need to install a vortex breaker for a nozzle or not, but generally speaking vortex breakers are not needed for huge tanks and may be needed for a small vessels only.

Down comers: in some cases we need to install down comers if we need to take liquid from a specific zone

of a container below the nozzle. It is important to know that the location of the nozzle and the location of the tip of down comer doesn't change anything about the hydraulics of a system. The hydraulics of a system is dictated by the pressure or liquid levels in the source and destination. Down comers can be used in some liquid containers with dirty service liquid. A down comer in this case helps us to remove all the sludges from the bottom of a container.

Riser: risers can be used as the internal portion of nozzles if we intend to remove fluid from a specific portion of a container above the nozzle. For example, one common application of risers is for overflow nozzles. Generally speaking overflow nozzles on containers should be placed at the top of the container at a point that is higher than the HHLL. However, this location is not the best location from an operations point of view. It is very high and could have limited accessibility. One option is installing overflow nozzles on the lower side of the container and putting an internal riser where its tip goes up to the HHLL.

Extended nozzle: nozzle extensions can be used for the cases where, for whatever reason, the incoming fluid is not supposed to be in contact with the container walls before getting enough homogeneity with the rest of fluid in the tank.

Elbowed nozzles are used for different reasons. It could be to dissipate the energy of the stream. The other reason could be to provide a good flow pattern in the container. Elbowed nozzles are not generally used for storage tanks.

The above are only a few types of nozzle internals. The nozzle internals could be more complicated systems too. One example is a nozzle with a floating pipe (Figure 9.20). This system guarantees always taking liquid from the surface of the liquid bulk.

9.9.7 Nozzle Externals

The nozzles on containers could be connected to long pipes or instruments. These are not generally considered as "nozzle externals" per se.

However, there are some cases where a short piece of pipe is connected to the nozzle from outside of the nozzle. One example is a gooseneck on vent nozzles.

9.10 Overflow Nozzles

Overflow nozzles are obviously only for containers with liquid in them. Overflow nozzles are basically "safety systems" to protect containers against filling out. Not every single container is equipped with an overflow nozzle. In fact, overflow nozzles are more common for tanks than vessels. However, even for some tanks the overflow nozzle is not considered! There could be cases that we cannot afford to see overflowing liquid outside of them. For example it could be the case that it is very risky for a very flammable liquid to overflow into the dyke around the tank because of the chance of fire. In such cases we may decide to overlook the overflow nozzle and instead implement a more reliable tripping system to protect the tank.

The convention and simplest types of overflow nozzle are shown in Figure 9.21a. Later it was found in operation that overflow systems create a lot splashing, which prevents operators from coming close to the overflowing tank. Then in the updated versions of overflow systems, a vertical pipe connected to the nozzle (Figure 9.21b). After a while they realized that in tanks that frequently overflowed, the overflow stream washes out of the tank foundation. An elbow is connected to the end to solve this problem (Figure 9.21c).

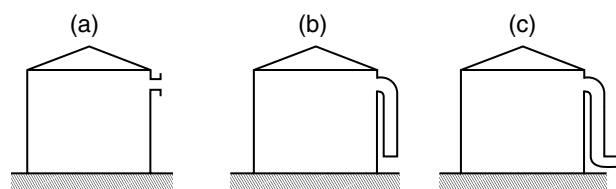
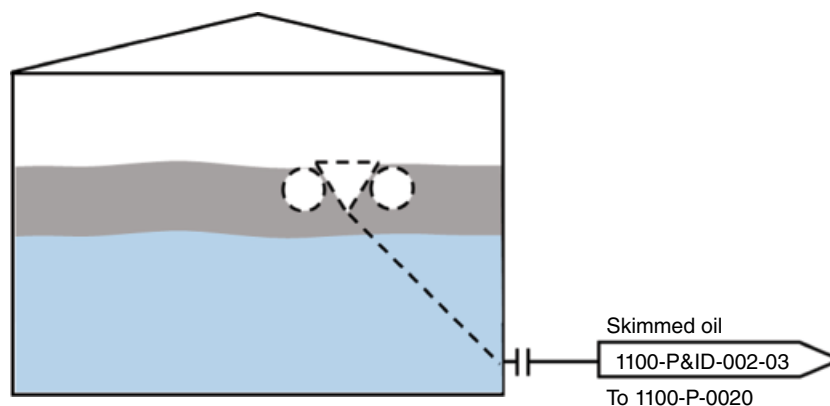


Figure 9.21 Conventional overflow system.

Figure 9.20 Nozzle with an internal floater.



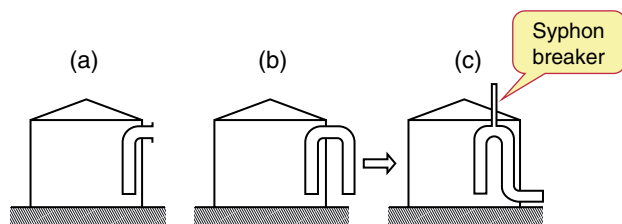


Figure 9.22 Inverted “U” overflow system.

The problem is that this overflow system doesn’t work for blanketed tanks. Blanketing systems will be discussed in Section 9.12, but it is basically a system that exerts a positive pressure on the surface of liquid in a container. The blanketing gas in a blanketed tank with a conventional overflow is swept away through the overflow nozzle. Then, for such cases we need more complicated overflow system. Figure 9.21 shows such an overflow system with the same historical improvement. However, a new problem is created in the latest version of this overflow system (Figure 9.22c). As can be seen it is basically an inverted “U” which works as a syphon. This means as soon as there is overflow stream through this system, the flow will continue even after decreasing the liquid level to the normal level. This is because of syphoning. Then, to “break the syphon” and prevent continuation of the overflow stream, a “syphon breaker” should be attached to the neck of the syphon to break the created vacuum in it (Figure 9.22c).

9.11 Breathing of Non-flooded Containers

Breathing, or in more complete term, normal breathing, is basically “respiration” or the attempt at respiration of containers. Containers, similar to humans, breathe by inhalation and exhalation!

Breathing means the generation and condensation of vapors existing on the liquid surface of non-flooded containers. The generation of vapors is called out-breathing and the condensation of vapors is called in-breathing.

Breathing is more important for tanks than vessels for different reasons. One is the fact that in tanks the surface area of the liquid (in contact with air) is higher than for vessels, and also the fact that tanks are less capable of handling high pressures than vessels. Therefore means should be implemented to take care of tank breathing.

The breathing could be thermal breathing or mechanical breathing. Mechanical breathing happens

because of inequality of flows around the container and thermal breathing because of ambient air temperature changes.

For example, in mechanical breathing, when the flow-in is higher than the flow-out, the liquid level in the tank goes up and it pushes the vapor to go out of the tank, or, if there is no means to handle that, the tank will be pressurized.

The concepts of breathing are shown in Table 9.10.

There are different ways to handle tank breathing.

One method of handling of tank breathing is “free vents.” Free vents are the simplest and the most common systems to take care of tank breathing. A free vent could be a nozzle on the top of the container with a simple, short piece of pipe connected to it. These are installed on the roofs of containers (tanks).

Table 9.11 summarizes the different free vent arrangements.

Sometimes the tank-sitted vents are unacceptable from a regulatory viewpoint and an elevated vent is needed. Cold vent stacks are used to emit the vapors at a higher elevation to satisfy the regulatory bodies (Figure 9.23).

These days free vents are less common than in older times. This is because of more stringent environmental regulations. If the vapors are not considered “innocent” by the regulatory bodies, they cannot be released to the atmosphere. Therefore some vapors may need to be treated before releasing to the atmosphere. The type of treatment depends on the type of vapor. The treatment could be condensation, adsorption, adsorption, drying, or even burning. “Vent scrubbers,” “vent condensers,” “vent adsorbers,” bin filters, and flares are a few of the vent treatment systems.

The treatment systems could be in situ or ex situ.

The in situ systems are the ones that are installed on the vent pipe of the tank, on roof tanks, or on the ground beside the tank to treat the vapors out of tanks.

The ex situ systems are central facilities that treat the vapors from several tanks, or the vapors can be sent to a remote central treatment system for ex situ treatment. The vapors from several tanks are collected through a vapor collection network and sent to the VRU. The ex situ treatment systems for vapor treatment are generally named “vapor recovery units” or VRUs (Figure 9.24).

In situ or ex situ vapor treatment systems are in different forms depending on the type of vapor. They could be scrubber type, condenser type, or absorption type.

If the release vapors to atmosphere are flammable a flame arrestor may also need to be installed.

Table 9.10 Concepts of container breathing.

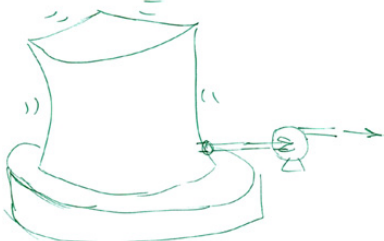
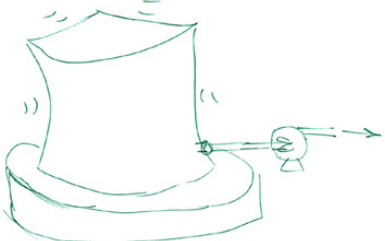
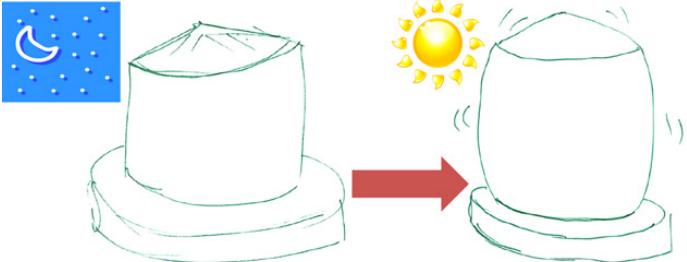
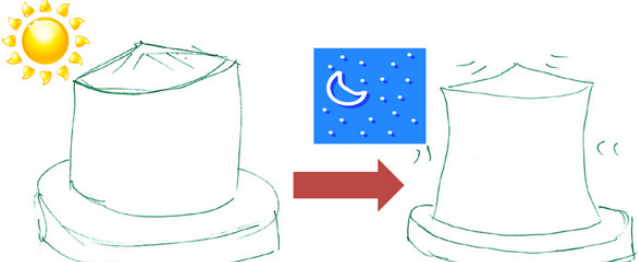
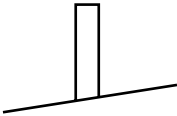
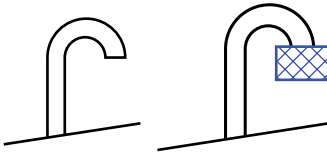
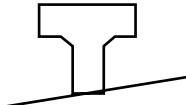
	Out-breathing	In-breathing
	Creates more vapors – increases the container pressure	Condenses vapors – generates a vacuum in the container
Mechanical Breathing		
Thermal breathing	<p>Because: flowing-in</p>  <p>Because: weather heating up</p>	<p>Because: flowing-out</p>  <p>Because: weather cooling down</p>

Table 9.11 Different types of free vents.

	Straight type	Gooseneck type	Ell type
Vent type			
Pros	The least expensive	Mid-price	Less troublesome
Cons	Risk of getting in the rain and debris	Risk of dripping of condensate on the roof	The most expensive

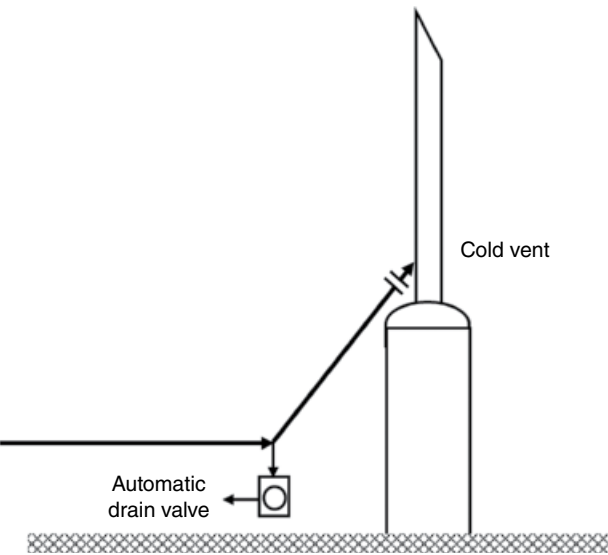


Figure 9.23 Cold vent stack.

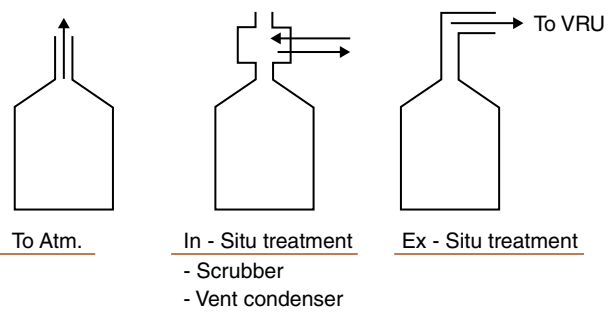


Figure 9.24 Different methods of handling vent vapors.

9.12 Blanketed Tanks

When the goal is to store volatile liquids in a container, a floating roof tank can be used. The other option is to use a fixed-roof tank with a blanking or padding system. Blanketed tanks can be used if the liquid of interest is volatile but not very volatile.

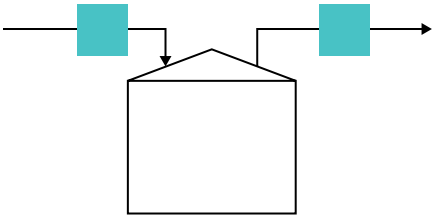


Figure 9.25 Blanketed tank.

A blanketing system provides a positive pressure over the surface of the liquid inside the tank in order to minimize the escape of volatile liquids, and also to provide a safe atmosphere inside the tank (Figure 9.25).

To implement a blanketing system two questions need to be answered: what type of blanketing gas should be used, and what type of blanketing system should be designed?

Blanketing gas should be an inert gas to minimize the chance of flammability and corrosion. The most common blanketing gases are nitrogen, natural gas, and carbon dioxide. Nitrogen is the best and the most expensive gas for blanketing purposes. However, we can't always afford to use nitrogen as the blanketing gas. The most common use of nitrogen gas as a blanketing gas is in the food industry, where the products are expensive and need to be in contact with a food-grade gas. In the oil industry, the most common type of blanketing gas is natural gas, which is mainly composed of methane. Methane can also be considered as an inert gas. You may be surprised that natural gas is "inert" since it is very flammable. However, the fact is that natural gas is only flammable when it is mixed with air. Natural gas, when used as a blanketing gas, is always enclosed in the top space of the tank and it is not in contact with air, so there is no chance of fire and it can be considered as an inert gas in this application. The last type of blanketing gas is carbon dioxide (CO₂). Carbon dioxide can form acidic vapors when it is in contact with water. These acidic vapors are corrosive. Therefore CO₂ is generally useable in applications where the equipment is already corrosion resistant.

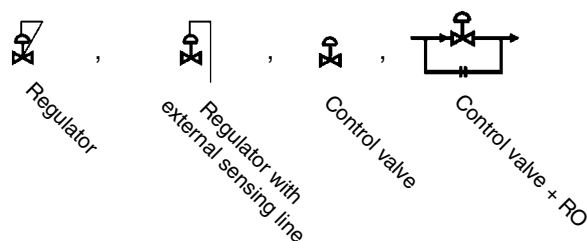


Figure 9.26 Different arrangements for regulating gas and vapor streams.

The major application of CO₂ as blanket gas is in the food and beverage industries.

To answer the second question it should be said that, generally, a blanketing system on a tank includes two nozzles and pipes connected to the roof of the tank: one pipe brings the blanketing gas into the vapor space of the tank and pressurizes that space up to a certain pressure, while the other pipe removes the blanket gas, which may be saturated with the vapor of the stored liquid. There needs to be a control system on each of these pipes, otherwise the blanket gas going into the tank quickly moves out of the tank and is wasted. The control system could be a control valve, a pressure regulator, or a “restriction orifice” (Figure 9.26).

There are different control systems on these two pipes to satisfy the requirement of the blanketing gas system; these will be discussed in Chapter 14.

9.13 Heating (or Cooling) in Containers

There could be cases where a container needs heating. However it should be mentioned that the purpose of putting any type of heating mechanism in containers is not generally to increase the temperature of the liquid. If there is an intention to increase the temperature of a liquid stream the best equipment is a heat exchanger. Heating in containers is not very efficient and we do it only if we have to. Generally speaking heating mechanisms in containers should be installed only for the purpose of temperature maintenance. This means we use the heating mechanisms just to keep the temperature of the containers content.

There are two main types of heating in containers: heating using heating fluid and heating using electrical heaters.

Heating using heating fluid can be done by coils of heating media. The heating media could be steam, hot glycol, etc. Heating coil can be laid down on the floor of the containers or, as can be seen in newer facilities, on the shell of the containers (Figure 9.27). The reason that companies try to not using heating coils on the floor of the containers is because of repeated breakage of them caused by operators walking over them during inspection

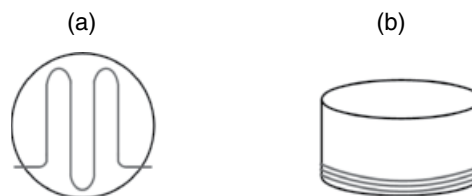


Figure 9.27 Two types of container heating by heating fluid.

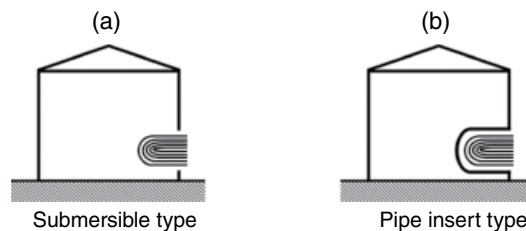


Figure 9.28 Tank electrical heaters.

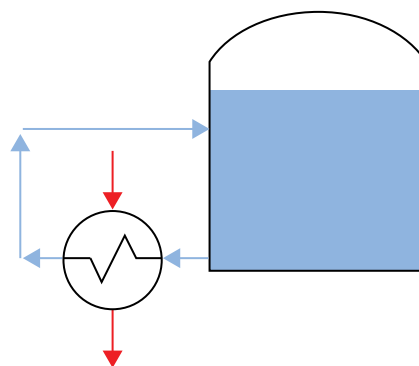


Figure 9.29 Complementary heat exchanger for a container.

and maintenance. The better version is putting the heating coils on the perimeter of the containers and up to a certain height of the containers. Generally there could be between two and five rounds of heating coils around a container. However, the exact number of rows is calculated during the design phase of the project.

Heating using electrical heaters is a very expensive solution. It could be done when heating using heating media is not applicable. Electrical heaters for tanks are mainly manufactured in two main types: submersible type and pipe-insert type (Figure 9.28).

The advantage of the pipe-insert type electrical heater is that the heater can be removed and be inspected during normal operation of the tank and without the need for emptying the tank.

As was stated, if more intensive heating is needed heat exchangers can be used. The heat exchanger could be placed in straight through system or recirculating system. The recirculating system is considered as a complementary heat exchanger to the containers (Figure 9.29).

There are some fire-based heating systems that are very common and are applicable in smaller size containers. For example fire-tube equipped containers are mainly applicable for vessels and not tanks. Submerged combustion is another method, but it is not very common.

Steam injection is a method that is sometimes used.

Heat tracing, heat blankets, and steam jackets are generally considered as winterization systems and not heating systems. However, they can also be used for heating small vessels. Winterization methods will be discussed in Chapter 17.

9.14 Mixing in Containers

Sometimes it is necessary to carry out mixing on streams after merging them together. If mixing can be done in a pipe only a static mixer is enough. However, the other type of mixing can happen in containers. There are two main types of mixing in containers: the hydraulic type and the mechanical type. Different types of mixing in containers are shown in Figure 9.30.

The mechanical type of mixing is basically using a mechanical mixer for the purpose of mixing. Mixers are a type of impeller that is connected to an electric motor. The types of mixer impellers are different and depend of the type of mixing, and are beyond the scope of this book. Electric motors could be connected to impellers directly or through gear boxes. However, we don't see this in the P&ID unless the gear box is huge and needs some sort of lubrication system. Mechanical mixers can be installed in two different positions in containers: on the top and center of the container and at the side of the container. Installing the mixer on the top of the container is a better option; however, it is not always applicable unless the diameter of the container is small (say <5 m in diameter). If the diameter of container is large as in tanks (say >5 m diameter) using a top mounted mechanical mixer is not applicable and a side mounted mixer or mixers should be used.

If the mechanical mixer is top mounted the best position is at the center of the container. Installing the mechanical mixer at the center of the container minimizes stress and vibration on the container. However, to make sure that mixing happens, rather than rotation of the whole bulk of fluid inside of the container, some baffles should be

installed on the container. If, for whatever reason (e.g. risk of plugging because of liquid dirtiness), installing baffles is not applicable we may choose to install a draft tube or off-center top mounted mechanical mixer. The other option is installing a top mounted, tilted mixer. However, the top mounted center mixers are the best choice.

For large tanks side mounted mixers are more common. Even though operators always have concerns about leakage from side mounted mixer nozzles, sometimes using them is inevitable. Side mounted mixers could be installed on the shell side of tanks from one to three or more units.

The second type of mixing in containers is hydraulic mixing. Hydraulic mixing is not a very efficient type of mixing in comparison to mechanical mixing. If, for whatever reason, mechanical mixing cannot be done, for example, when the fluid is very aggressive against the mixer impellor, hydraulic mixing can be done. Hydraulic mixing can be in the form of recirculation or direct injection. Recirculation means sending back a portion of liquid from the discharge side of the pump back to the tank, but in direct injection just the incoming fluid goes through the nozzle and through the sparger in the tank. The recirculation pipe is connected to a nozzle on the tank and the nozzle internally is connected to a piece of pipe in simple form or in the conductor type.

9.15 Container Internals

The necessity for and type of container internals are specified during the design of equipment. However it should be mentioned here that the tank internals are never a welcomed item! We put them if we have to. Internals in containers are very difficult for inspection, and if they are malfunctioning it is not easy to spot them.

We do not always show the container internals on P&IDs, but if we choose to show them (because of their importance) we show them in the form of dashed lines.

9.16 Tank Roofs

There are several types of tank roofs but the one that can be recognized in P&IDs are dome roofs, fixed cone roofs, fixed cone-internal floating roofs, and external floating roofs.

Fixed roof tanks are for non-volatile liquids.

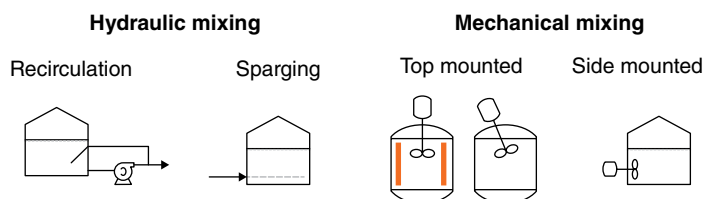


Figure 9.30 Different types of mixing in containers.

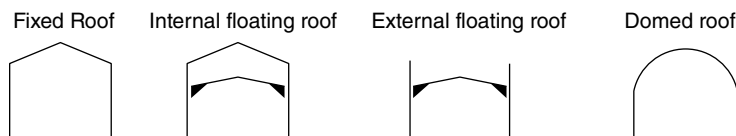


Figure 9.31 Tank roofs.

Floating roofs were invented to store volatile liquids. The first type of floating roof was the type later called “external floating roofs.” The external floating roofs have some fugitive gases that then pollute the air. After environmental regulations became stricter a fixed roof was installed on the floating roof and “internal floating roof tanks” came on to the scene.

Dome roof tanks are the tanks that have a fixed roof but the type of their roof is not cone type but dome type.

Figure 9.31 shows different types of cylindrical tank roofs on P&IDs.

9.17 Tank Floors

Floors of tanks can be of two main types: flat floors and sloped floors (Figure 9.32).

From a construction and also capital cost viewpoint the best tank floor is a flat one and is the easiest floor to construct. However, in some cases, because of operational reasons, the floor of a tank may need to be constructed in a sloped form.

There are mainly two types of sloped floors: radially sloped floors (double sloped) and diametrically sloped floors (single sloped).

Radially sloped floors are of two types: cone down and cone up. Radially sloped floors are much cheaper than diametrically sloped floors. In fact, diametrically sloped floors are not very common in the industry for different reasons, including their high cost of construction. Diametrically sloped floor tanks are not very common for tanks with diameters more than 20–30 m.

The sloped floors allow easy material movement within a tank and eventually easy removal of liquids or

flowable solids. For example, silos almost always have a sloped floor to facilitate the removal of solids from the silo to the outlet nozzle. The type of sloped floor in silos is the coned-on radial type. For complete drainage of a tank we need to have a sloped floor. This is especially important when the liquid content of the tank is flammable or toxic. In such cases before an operator enters for inspection or maintenance the tank should be completely drained. In that case the tank may have a coned-down floor or a cone-up floor.

One important thing that is sometimes overlooked is that even flat floor tanks will turn to cone-up floor tanks after a while. The reason is if the weight of the shells on the tank is larger than the weight of floor sheets then the shells push the perimeter of the tank down and consequently the center of the floor goes up.

If the intention of a sloped floor is complete drainage the drain valves should be on the side of liquid accumulation. This means that in a coned-down arrangement the drain valves should be extended to the center of the tank inside of a sump. If the floor arrangement is cone up there should be multiple drain valves around the perimeter of the tank (Figure 9.33).

Arguably the most common tank floors after the flat floor is the cone-down type. Here we are going to shed more light on the angle of tank floors in the cone-down cases.

The angle of cone-down floors could be in a wide range. It may start from less than 1° to a large value of 60° . For liquid in tanks, the floor cone-down angle could be a number from 1° to 5° for low viscosity liquids to a number from 10° to 15° for high viscosity liquids. For flowable solids in silos, the floor cone-down angle could be a number from 45° to 60° . Interestingly the floor cone-down angle would be a number around 0.5° if the plan is to have a flat floor

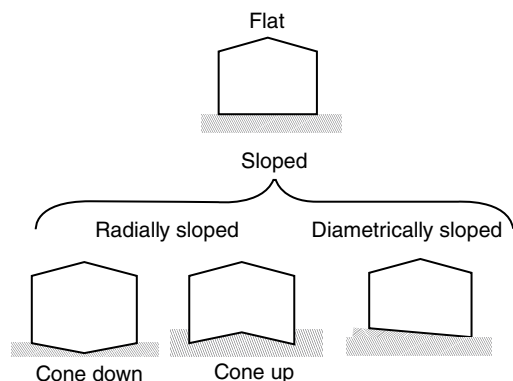


Figure 9.32 Tank floors.

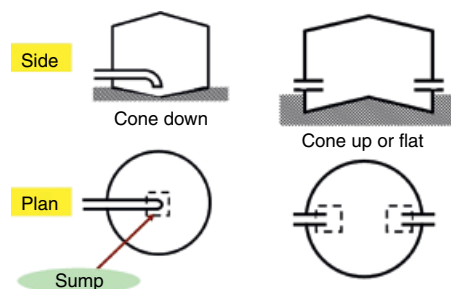


Figure 9.33 Drain valve arrangement for cone-down and cone-up floors.

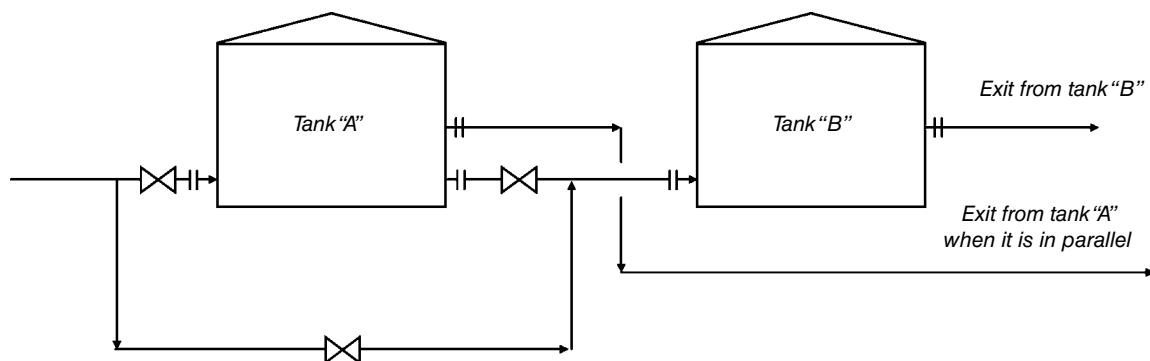


Figure 9.34 Tanks in parallel and series arrangements.

and it is coned because of the higher weight of the tank walls, as was mentioned before. The selection of floor cone-down angle is a process decision to facilitate material movement toward the center. The floor cone-down angle could be limited because of construction implications. Generally speaking wherever the tank has a larger diameter, it is more expensive to have a larger floor cone-down angle. For large diameter tanks (possibly more than 5–10 m diameter) the floor cone-down angle is generally kept below 5°. A rule of thumb that can give a guideline for the maximum allowable floor cone-down angle – from a constructability viewpoint – is as follows:

$$S_{\max.} D = 30$$

where:

$S_{\max.}$ is the maximum allowable floor cone-down angle in degrees (°) and D is the tank diameter in meters (m).

However, there are some cases where the tank has a large diameter and for process reasons it should be high angled. There are some solutions for that. For example in some clarifiers or thickeners the floor should be high to help directing the settled solids (sludge) toward the center and finally removing it. If the clarifier or thickener is a large diameter one, the slope is possibly limited to 5° or a value around that. A rake system can be implemented to sweep the settled sludge toward the center of the tank.

9.18 Container Arrangement

Containers could be in a series and/or parallel arrangement when they are used for unit operation or process units. As tanks are less likely used for unit operation or as process units, vessels are more commonly seen on P&IDs in series or parallel arrangements.

If a vessel is supposed to be used for holding (short time) it is usually a single vessel.

If tank(s) are supposed to be used for storing material (long term) it could be single in the majority of cases.

However, they could be in multiple arrangements. This multiple arrangement could be neither parallel nor in series. The tanks could be connected to each other with a specific piping arrangement to provide flexibility for the operators to use them in series or parallel, depending on the case. Figure 9.34 shows such an arrangement.

9.19 Merging Containers

Sometimes containers are merged together to save money. There are at least three ways to merge containers: complete merge, merge with volume dedication, and merge while keeping physical boundaries.

Complete merge is using one container for more than one purpose. This can be done by replacing two or more containers with only one of the same or larger volume.

“Merge with volume dedication” has the same concept as “complete merge” with only one difference. In this type of merge the volume of the container is “somehow” dedicated to each user or purpose.

For example in Figure 9.35 a single tank is used for fire water AND plant water. By a specific internal arrangement the plant water is allowed only to use the

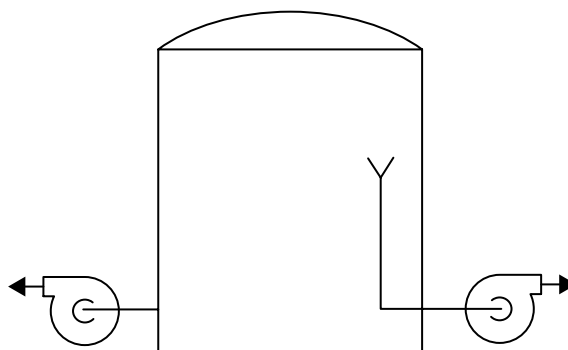
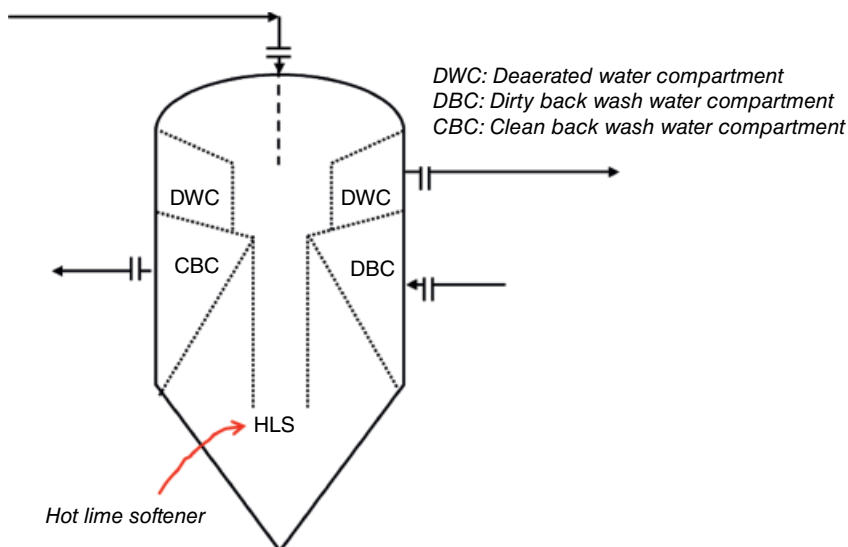


Figure 9.35 Dedication of tank water to fire water and to plant water.

Figure 9.36 Merging tanks into a compartmented tank: HLS example.



top portion of the tank volume. The rest of the tank volume is dedicated to fire water.

The third type of merge is a container with internal compartments. One good example of this type of merge is in hot lime softener (HLS) in water treatment areas. HLS is an expensive piece of equipment that, in addition to other features, should be installed on legs as it has a sloped floor. This makes the installation of HLSs expensive. This has made companies think of using HLS internal space for other uses. For example a compartment could be fabricated inside of HLS and be used as a “backwash water tank” for filters downstream of the HLS (Figure 9.36).

9.20 Secondary Containment

Secondary containment is a physical enclosure around voluminous items to prevent wild liquid escape during an uncontrolled (i.e. accidental) release. The requirements for secondary containment could be technical and/or legal. A regulatory body in a specific jurisdiction may ask for secondary containment to be provided for all tanks and/or vessels and/or pipes in a plant.

Secondary containment is used as a safety measure just in case there is a sudden rupture in the body of a large tank to prevent the release of a huge amount of liquid to the plant and the neighborhood. Some may think that secondary containment is only necessary when a tank or tanks contain “non-innocent” (non-harmful) liquids. It was the case in the old days that secondary containment was only used for non-innocent liquids. However, these days secondary containment is applicable for all type of liquid content, even potable water. However, the local codes define whether you need to put secondary containment on a tank or not.

The other issue is if secondary containment is not only for tanks. Even though it is very common to see secondary containment for tanks, secondary containment could be done for tanks, vessels, or even pipes. In some process plants where they produce lethal material they may have secondary containment on their tanks and vessels, and even on the pipes.

You may have seen secondary containment in process plants around tanks or a group of tanks. They are generally in the form of dykes or berms.

There are two main methods to provide secondary containment for voluminous elements:

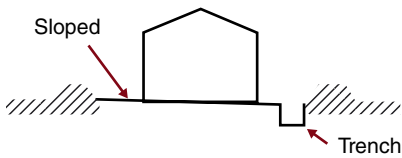
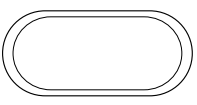

- 1) Providing a berm or dike around the element
- 2) Using a double wall element.

In both the above methods we have primary and secondary containment. The wall of the container or pipe is named the primary containment and the second barrier that we put around the container or pipe for safety purposes is the secondary containment.

These two methods are shown in Table 9.12.

There is one critical characteristic for secondary containment that, if it fails, the physical containment cannot be qualified as secondary containment, which is the space between the primary and secondary containments. The important point is that there should be a space between the primary and secondary containments in a form that the space can be monitored against the leakage. For example, for very common dykes around tanks the space between the dyke and the tank body is visible to the field operator. The field or rounding operator can always check if there is a leakage from the tank to the dyke area and warn the operating personnel before a large rupture and large release of liquid.

Table 9.12 Concept of secondary containment in containers and pipes.

	Primary containment	Secondary containment
Tank/ vessel	Tank wall	Dike: 
		Double wall container 
Pipe	Pipe wall	Double wall pipe: 

For double wall containers or pipes such monitoring can be done in other ways. For example, in one way the interstitial space in vacuumed and the vacuum level is monitored by a pressure gauge (Figure 9.37a).

The other way is sort of blanketing the interstitial space and monitoring the differential pressure (Figure 9.37b).

9.21 Underground Storage Tanks

In today’s ever tightening environmental standards, underground tanks (or vessels) are not a favorite option.

Several leakages of liquid from underground storage tanks led regulatory bodies to limit the usage of them. One still acceptable application of underground storage tanks is gasoline storage in gas stations.

As was mentioned before, the secondary containment of underground tanks can be provided by using double wall vessels.

The other option that provides secondary containment is using in-the-pit vessels instead of underground tanks.

Although an underground storage tank is a buried vessel its nozzles and valves and its connecting pipes should be accessible. The nozzles should be placed on the top of the vessel and appropriately long nozzle necks provide

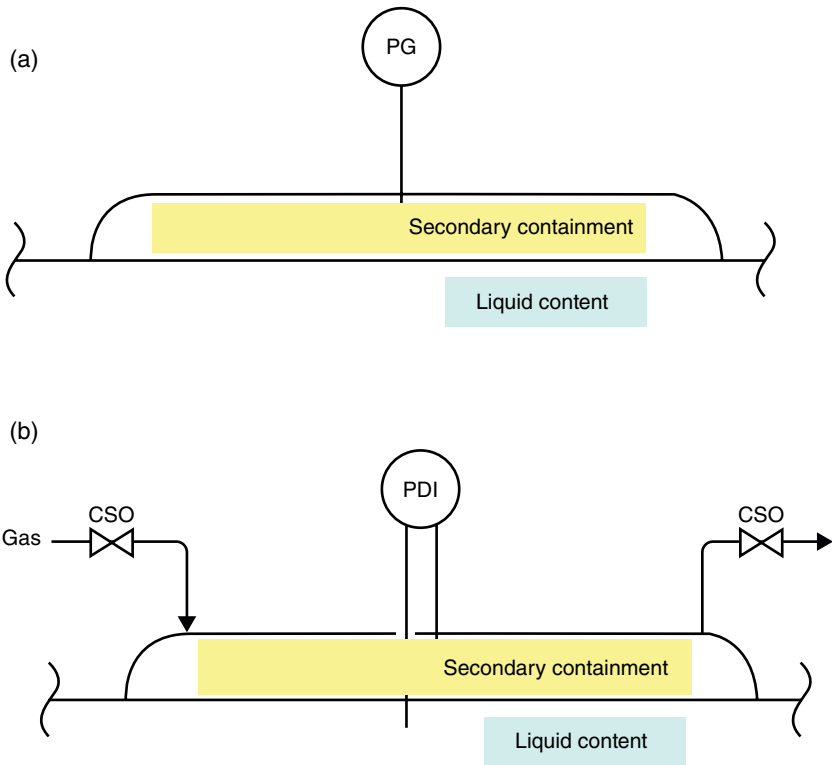


Figure 9.37 Leak monitoring of interstitial space in double wall containers.

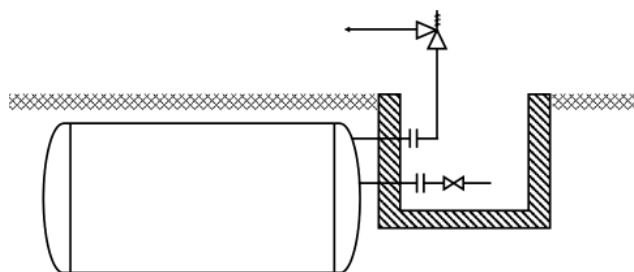


Figure 9.38 Underground vessel with dry vault.

the opportunity to have the nozzle flanges above ground level and accessible. The valves around the underground tank could be placed in a “dry vault” (Figure 9.38).

9.22 Sumps

Sumps are underground or in-the-pit vessels or basins for the collection of surface drainage. There are at least four types of surface drainage: storm water, floor washing water, drainage liquid from the draining operation of equipment, and discharged fire water.

Because all these water streams are running on the floor of a plant and don’t have enough pressure, they only can be collected to a lower elevation container. There, water is directed toward the sump through suitable sloping floors and/or trenches (Figure 9.39).

The sumps are primarily considered to collect watery liquids. Such liquids may have other constituents such as oil, chemicals, suspended solids, etc. If the amount of non-water constituents or aggressive chemicals is high, it may be required that the stream is directed to a dedicated sump. This sump could be called a chemical sump or oily water sump.

Sumps were originally built in concrete. However, because of more stringent environmental regulations regarding underground water contamination, metallic or plastic containers in the form of single walls (conventional) or double walls are more and more in use.

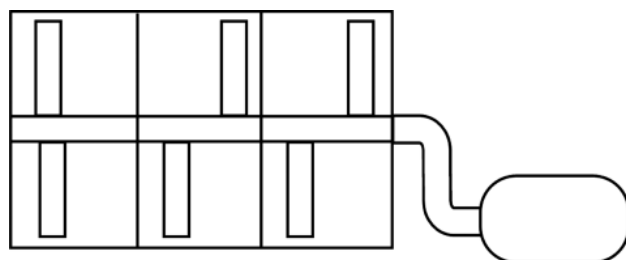


Figure 9.39 Plan view of a sump system.

Each sump has one, two, or more pumps to transfer the collected water to its destination.

The capacity of sumps is designed during the sizing phase of the project.

9.23 Wrapping-up: Addressing the Requirements of the Container During its Lifespan

In this part we check our design to make sure we have covered all the needs of containers during the different phases of the plant. As it was discussed before these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item.

The requirements are listed here:

- 1) Normal operation.
- 2) Non-routine operation (reduced capacity operation, start-up operation, upset operation, planned shutdown, emergency shutdown). These phases of the plant need two components to be handled: process components and control/instrumentation components. The process component is discussed here but the control/instrumentation component will be discussed in Chapters 13, 14, and 15.
- 3) Inspection and maintenance. General consideration regarding inspection and maintenance of all items is covered in Chapter 8. As containers could be so large to be walk-in, the isolation and cleaning of them should be taken very seriously.
- 4) Operability in the absence of one item. Generally no spare is provided for containers if they are for storing or holding purposes. This is especially true for tanks that are very expensive process items. If for whatever reason a tank goes out of operation the flow may need to be redirected to an external container or pond, or the container should be bypassed. Because of this it is better to try to design and install a few smaller tanks rather than one large tank. It has been seen in some cases that one tank is provided as spare for several other tanks. In such cases appropriate piping should be provided between the tanks. If it is intended to bypass a container in the case of absence, it should be considered during the design stage of a project to make sure the stream has enough pressure to get to the new destination downstream of the container of interest.

10

Pumps and Compressors

10.1 Introduction

Here the phrase “fluid movers” is used as a collective term for pumps and compressors. Note that this name is not common; it is only used here for educational purposes.

Fluid movers could be “liquid movers” (pumps), or “gas movers” (compressors, fans, and blowers).

10.2 Fluid Mover Roles

There are two possible reasons for using a fluid mover: transfer (or mobilization), and pressurization.

Liquid movers (pumps) are only used for transferring liquids, while gas movers could be used for mobilization only, or for both mobilization AND pressurization.

Compressors are gas movers which are used for gas mobilization AND pressurization. Fans, on the other hand, are used only for gas mobilization. Blowers could also be considered as gas movers that have the combined duties of mobilization and pressurization.

This concept can be seen in Table 10.1.

It is important to note that when using a pump the fluid should be only liquid; pumps have a very low tolerance for gas bubbles in the liquid. In the case of compressors, the fluid should be gas; compressors have a very low tolerance for liquid droplets in gas streams. That is why when a liquid is near its boiling point a specific control system must be implemented to make sure that no “vapor” will be formed and get into the pump. Similarly, in the case of compressors, if there is a chance that the vapor stream contains liquid droplets, they must be removed upstream of the compressor.

10.3 Types of Fluid Movers

Both pumps and compressors/blowers/fans are designed using two main mechanisms: dynamic type and positive displacement (PD) type. In the dynamic type, movement

of fluid is achieved by “throwing out” the fluid, while in the PD type, movement of fluid is achieved by “passing along pockets of fluid.”

Dynamic type fluid movers can be either axial type or centrifugal type. Positive displacement type fluid movers can be either rotary type or reciprocating type

Axial type fluid movers are mainly used for high flow rates and low-pressure systems. Positive displacement types are generally used for cases in which the flow rate is not very high and high pressure is needed at the outlet of the fluid mover. This concept can be seen in Table 10.2.

In the table, capacity and differential pressure of fluid movers are indicated through a scale system of 1–4; 1 is the highest value of parameter and 4 is the lowest.

Axial type fluid movers are not very common unless a huge flow rate is needed.

The most common pumps in industry are the centrifugal type while the most common gas movers are PD types.

10.4 A Brief Discussion on the Function of Fluid Movers in a System

Fluid movers have two main parameters, which are used to specify fluid movers:

- 1) Flow rate or capacity (Q)
- 2) Differential head or differential pressure (ΔP or ΔH).

So, for example, if I want to buy a dynamic pump (like a centrifugal pump), should I say, “Please give me a centrifugal pump with a rated flow rate of $200 \text{ m}^3 \text{ h}^{-1}$ and a (differential) pressure of 500 kPa”?

You can do this, but what the vendor will give you is a pump with a maximum flow rate (rated capacity) of $200 \text{ m}^3 \text{ h}^{-1}$ and best efficiency at a (differential) pressure of 500 kPa.

Table 10.1 The varied traits of liquid versus gas movers.

	Transfer (mobilization)	Transfer and pressurization	
Liquid movers	Pumps	Not applicable	
Gas movers	Fans	Blowers	Compressors

The best efficiency (differential) pressure is the differential pressure that a centrifugal pump can generate while it is operating at its highest efficiency. In workplace conversations, the phrase “differential pressure” of a pump is used instead of the longer phrase of “best efficiency (differential) pressure.”

There is a different story for PD pumps. Let’s say you want to buy a PD pump with a maximum flow rate (rated capacity) of 20 gpm and a (differential) pressure of 300 psi. The vendor will give you a PD pump with a maximum flow rate (rated capacity) of 20 gpm and a maximum attainable (differential) pressure of 300 psi, which won’t explode the pump casing. The reason for potential explosions in PD fluid movers will be explained in the following paragraphs.

Table 10.3 summarizes the different meanings of “rated differential pressure” in two different types of fluid movers: dynamic and PD types.

So, if this is the case, then what would be the “actual” differential pressure of the pump I purchased?

The answer looks strange: The “actual” differential pressure is dictated by the system you use your fluid mover in!

If the fluid mover is a dynamic type (e.g. centrifugal type), the fluid mover installed will generate only enough discharge pressure to overcome the system resistance (meaning the destination pressure plus pipe friction resistance of the system), which could be a differential pressure lower or higher than the “rated” differential pressure or “best efficiency DP.” This means this fluid mover won’t necessarily operate at its optimum point (the point at which the fluid mover has the highest efficiency).

An axial fluid mover that doesn’t work on best efficiency points (BEPs) wastes a lot of energy and often experiences premature failure because of severe vibration.

For PD type fluid movers, the story is different. By installing a PD type fluid mover in a system, it will also generate only enough discharge pressure to overcome the system resistance, which could be a differential pressure lower than the “rated” differential pressure, and the fluid mover works with the same high efficiency! This means there is no “best efficiency DP” for PD fluid movers; all PDs can operate well at a DP lower than rated. The only “bad DP” is a DP that would cause the fluid mover casing to explode because of the high pressure generated.

Therefore, it can be seen that PD type fluid movers are compatible with all systems as long as their flow rate matches the system flow rate. However, installing an expensive high differential pressure PD type fluid mover in a low-pressure system is not an economical decision.

To summarize, it is important to align both the capacity and the differential pressure of a fluid mover to the system, disregarding the aligning of the differential pressure for axial type fluid movers creates technical problems, and disregarding the aligning of the differential pressure for PD type fluid movers may create an economical issue.

Table 10.3 Comparison of fluid movers and their differing definitions of rated differential pressure.

	Dynamic type	Positive displacement type
Rated capacity	Maximum flow rate the equipment can handle	Maximum flow rate the equipment can handle
Rated differential pressure	The differential pressure the fluid mover can generate when it is operating at its highest efficiency (best efficiency point: BEP)	The differential pressure the fluid mover can generate with enough safety margin to avoid explosion

Table 10.2 Comparison of different fluid movers.

Sub-type	Dynamic type		Positive displacement type	
	Axial type	Centrifugal type	Rotary type	Reciprocating type
Capacity (flow rate)	1 (highest)	2	4 (lowest)	3
Differential Pressure	4 (lowest)	3	1 (highest)	2
Popularity	Not common in pumps Not common in gas movers except for fans	Most common type of pumps (more than 80% of pumps)	Most common types of gas movers	

One main point of the above discussion is that specifying a pump as a “pump with $100 \text{ m}^3 \text{ h}^{-1}$ and a differential pressure of 200 kPa ” is not ideal. Each pump can work over a wide range of operating points, but there is a small operating window in which they work best from a technical and/or economical standpoint.

The other aspect of this concept is the example below. If you have a compressor with a capacity of $100 \text{ m}^3 \text{ h}^{-1}$ operating in a system that requires a differential pressure of 200 kPa , and you remove this compressor from the current system and try to reuse it in another part of the plant with roughly the same flow rate, the new differential pressure of the compressor in the new position could be different from than in the old position!

To explain this another way, each fluid mover operates not just at one point, but on a curve, which is called the “pump or compressor operating curve.”

This fact shows that there is a need for a control system to bring the operating point of the pump to the best point on its curve. Without any control system, a fluid mover “runs over the curve” away from the high efficiency point and without any limitation, and this is not good operation.

This concept is shown in Table 10.4.

What happens at the endpoints of the fluid mover operating curve? These points are explained in Table 10.5.

There is one point regarding the unit for differential pressure of dynamic fluid movers. Dynamic fluid movers transfer fluids by throwing out packets of fluid and this

type of fluid transfer gives them a specific feature: dynamic fluid movers generate a specific discharge pressure irrespective of the density of the fluid. For example, if a dynamic fluid mover can throw water a distance of 2 m , then it can do the same with liquid mercury, which is a heavy liquid. Because of this, instead of reporting differential pressure (in psi, kPa or bar) for centrifugal pumps, it is more common to report the injected energy to the fluid not in pressure unit as DP but in length unit as head.

10.5 Fluid Mover Identifiers

Based on the concepts stated in Chapter 4, the identifiers of fluid movers are symbol, tag, and call-out.

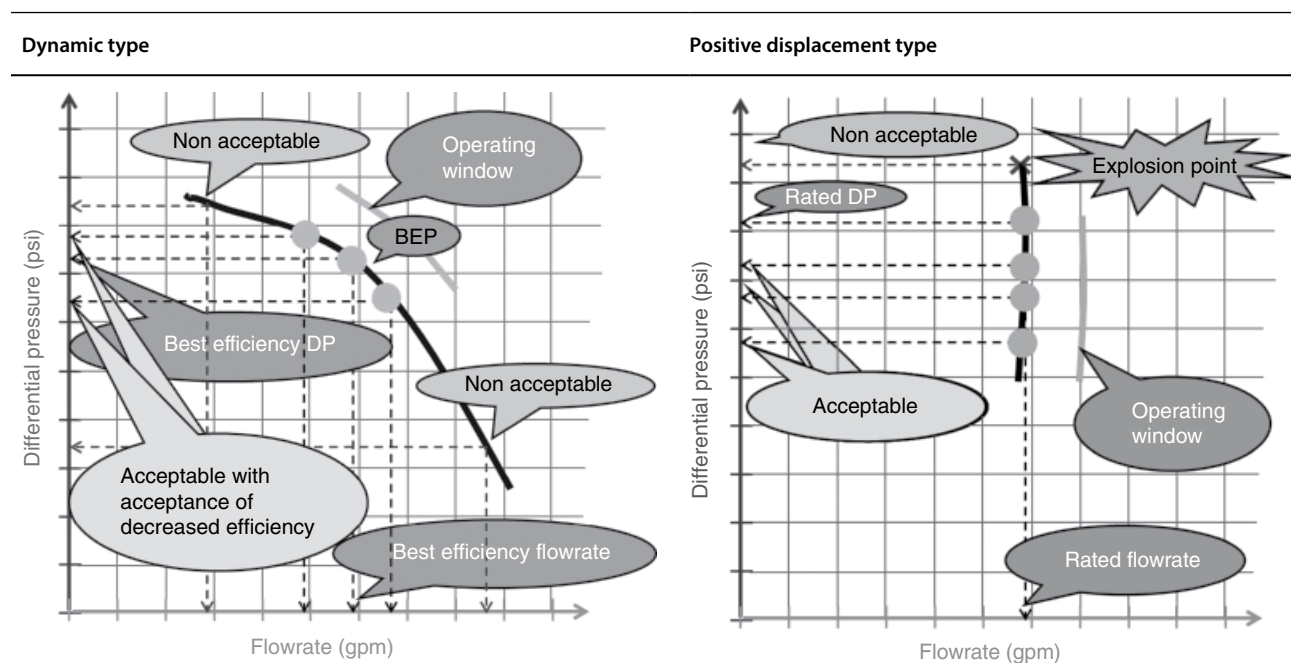
10.5.1 Fluid Mover Symbol

There are plenty of different symbols for various types of fluid movers in different companies. Table 10.6 shows some of them.

10.5.2 Fluid Mover Tag

The necessity of putting fluid mover tags on the body of the P&ID is mentioned in the project documents. If fluid movers tags need to be shown on the body of P&IDs they are generally placed below the fluid movers.

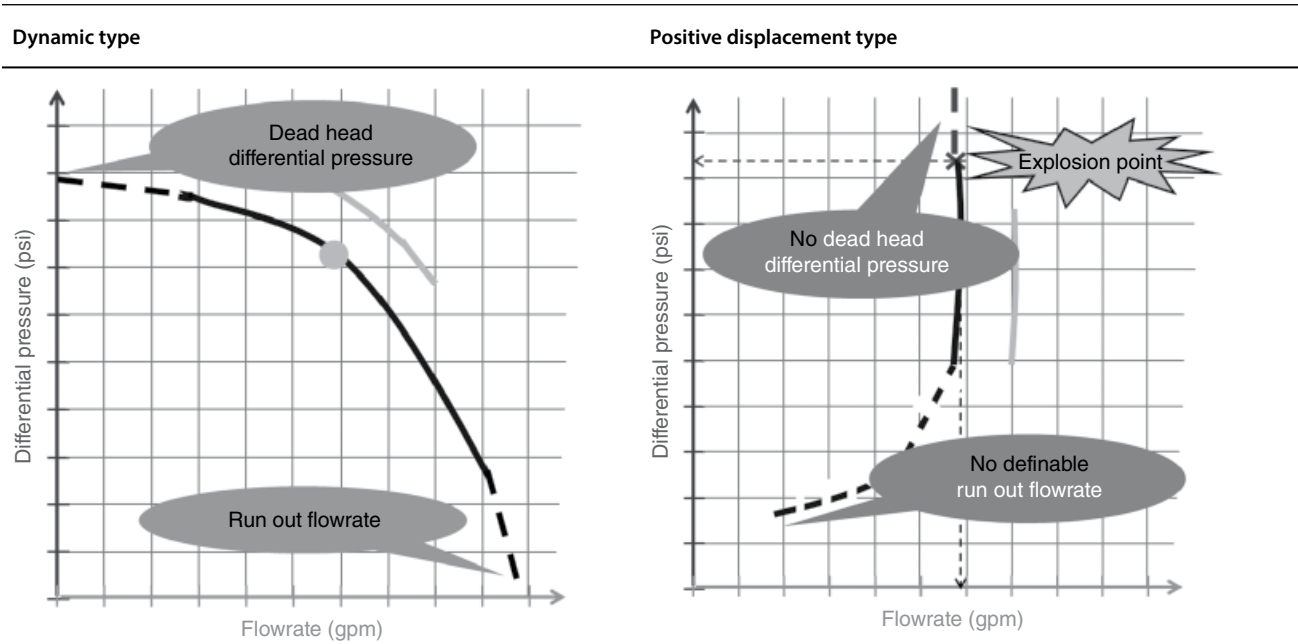
Table 10.4 Demonstration of the pump/compressor operating curve in differing types of fluid movers.



Summary: the operating curve of a dynamic fluid mover is “dropping and concave.”

Summary: the operating curve of a PD fluid mover is fairly vertical.

Table 10.5 Continuation of the demonstration of the pump/compressor operating curve in fluid movers.

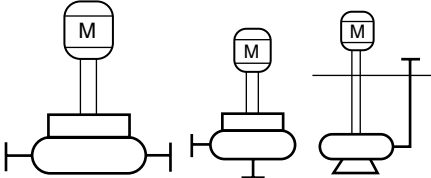
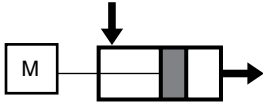
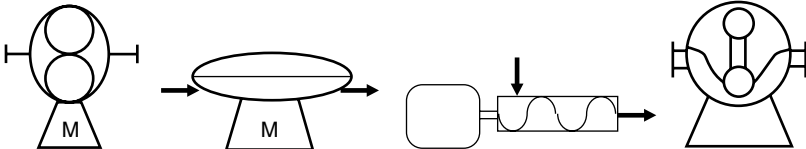


- Summary

 - 1) When the resistance against flow on the discharge side of a fluid mover approaches infinity, the fluid mover's operating point moves toward the left-hand side of the curve up to the "zero flow" point. This is the maximum pressure that the fluid mover can produce, but this pressure is not a practical number because it is generated when the fluid mover cannot move any flow (zero flow). The term for this highest static pressure is "dead head pressure." This situation happens when someone accidentally fully closes a valve on the discharge side of the fluid mover. The other term for this pressure is the "shut-off pressure."
 - 2) When the resistance against flow on the discharge side of a fluid mover approaches zero, the fluid mover's operating point moves toward the right-hand side of the curve down to the "zero pressure" point. This is the maximum flow rate that the fluid mover can move, but this flow rate is not a practical number because it is generated with zero pressure, which means non-moving flow! The term for this highest flow rate is the "run-out flow rate." This situation happens when someone accidentally detaches the outlet pipe from the discharge side of the fluid mover.
- Summary

 - 1) When the resistance against flow on the discharge side of a fluid mover approaches infinity, the fluid mover's operating point moves vertically upward but flow does not decrease (by much). Therefore, there is no "dead head pressure" that the system can experience, and if pressure increases to a value higher than "design pressure" of the system, the PD fluid mover and/or attached pipes will explode!
To avoid this, whenever there is a PD fluid mover, there should be a "relief device" in place (such as a pressure relief valve) to protect the system.
 - 2) There is no definable run-out pressure.

Table 10.6 Fluid mover P&ID symbols.

Type	Symbol
Centrifugal	
Reciprocating	
Rotary	

As it was mentioned in Chapter 4, the pump tag could be “P-3420” or “PU-3420.”

10.5.3 Fluid Mover Call-out

The typical call-out for a liquid mover is seen in Figure 10.1.

- The first line is the tag number of the fluid mover.
- The second line is the name of fluid mover.
- The third to fifth lines are the main characteristics of the fluid mover, which are flow rate (or rated capacity) and differential head (or differential pressure), or suction and discharge pressure separately.
- The next line is the driver type (if it is an electric motor, steam turbine, diesel-driven motor, etc.) and its power.
- The next line is the material of construction of the fluid mover; casing and internals (like impeller if the pump is centrifugal type).

Then we have minimum metal design temperature (MDMT), insulation, and trim.

Now let's discuss different types of pumps and compressors individually. Liquid movers (pumps) are discussed in Sections 10.6 and 10.7 but gas movers will be discussed in Section 10.8.

400-P-043
 WATER RETURN PUMP
 RATED CAPACITY: 166m³/h
 DISCHARGE PRESSURE: 1,600 KpAG
 DIFFERENTIAL PRESSURE: 1,500 kPag
 DRIVER: 224 KW @ 1,750 RPM
 MATERIAL: CASING-CS
 INTERNALS-12Cr
 MDMT: -29°C
 INSULATION: ET/10
 TRIM: ABA

Figure 10.1 A pump call-out.

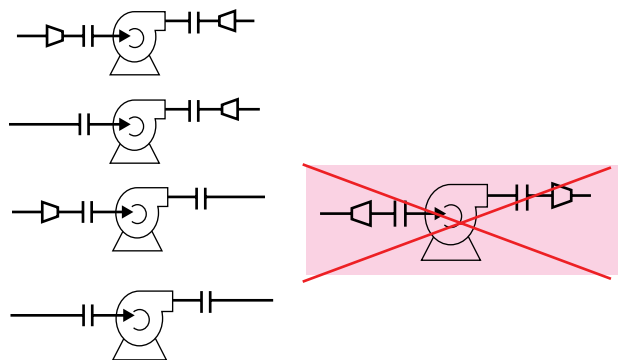


Figure 10.2 Various acceptable reducer and enlarger pairs.

10.6 Liquid Movers: Dynamic Pumps

Dynamic pumps are available for medium to high flow rates and low discharge pressure applications.

Centrifugal pumps are the most common type of pump in industry. Over 70% of pumps in industry are centrifugal. Centrifugal pumps are very operator friendly and operators usually prefer to work with them over PD pumps.

Axial pumps are not very popular pumps, and they have very limited applications. Sometimes these pumps are used for an application requiring the possible smallest shear exerted on the pumping liquid. One example is the “fish friendly axial pumps” for pump station applications.

In the discussion below one symbol (which is actually a PFD symbol) is used as being representative of centrifugal pumps and axial pumps.

10.6.1 Centrifugal Pumps

The piping arrangements for fluid movers could be most complicated piping arrangement around a general equipment.

The discharge flange of centrifugal pumps is generally smaller than the suction flange. In smaller centrifugal pumps the size of the discharge flange could be the same as suctions. Also, in some specific types of centrifugal pumps the size of discharge and suction piping are the same. One example is a regenerative turbine pump.

The flanges of centrifugal pumps are not necessarily the same size as their mating piping flange. It is very common to see that the piping flange on suction side of a centrifugal pump is different from the suction flange size and/or the discharge piping side of a centrifugal pump has a size different from the discharge flange of the pump. Therefore it is very common to see and to use reducers or enlargers to connect piping to centrifugal pumps. Different arrangements of reducer/enlarger pairs could be used on the suction and discharge side of centrifugal pumps and are all acceptable except in two cases: on the suction side of a centrifugal pump we never ever use an enlarger, and on the discharge side of a centrifugal pump we never ever use a reducer.

All other arrangements that are acceptable can be seen in Figure 10.2.

If, for whatever reason, we see that we have to use an enlarger on the suction side and/or reducer on the discharge side of a centrifugal pump there could be a mistake in design; in the design of the sizing of the pipe or of the sizing of the pump. It is important to know that there is not always a need for a reducer or enlarger on the suction or discharge side of a centrifugal pump. There

are some specific types of centrifugal pumps that don't need a reducer or enlarger. The other case for which we may not need a reducer or enlarger is for small centrifugal pumps.

Even though we talked about the potential necessity for a reducer/enlarger on either or both sides of centrifugal pumps, the real necessity and the required size of reducer/enlarger can only be recognized after receiving the pump vendor documents.

Sometimes during the early stages of P&ID development we put a reducer on the suction side of a pump and an enlarger on the discharge side of the pump, but without any sizing. However, later, after receiving vendor information, we can put the required size of reducer and enlarger, or we may need to totally eliminate them from the P&ID.

Field monitoring instruments for centrifugal pumps will be discussed in Chapter 14.

10.6.1.1 P&ID Development on the Suction Side

Reducers on the suction side of centrifugal pumps are more important than an enlarger on the discharge side. The reason is that in general centrifugal pumps the pipe on the suction side is usually a piece of horizontal pipe. When it comes to selecting a reducer to be installed in the horizontal position the question arises is whether this reducer should be concentric or eccentric. As it was mentioned before in Chapter 6, the less expensive reducer is always concentric type and this would be our default choice. However, there are cases where we have to use an eccentric reducer in the horizontal suction section of a centrifugal pump.

Using an eccentric suction reducer could be necessary if the flow is two-phase flow or "potentially" could be two-phase at the suction of the pump. A centrifugal pump doesn't have any problem with two-phase flow as long as the dispersed phase occupies a small portion of the flow. For example in liquid-gas flow, as long as the gas flow is fully dispersed and its content is low (say less than 5%) there is no problem in performance of a centrifugal pump. The other example is when there is liquid-solid two-phase flow. As long as the solid portion is fully dispersed and its content is low it's not harmful for a conventional centrifugal pump. However, the problem is when/if the dispersed phase is not fully dispersed. For example instead of having a dispersed gas or vapor bubbles in a liquid phase there is a packet of gas or vapor it would be detrimental for the centrifugal pump. If instead of a fully dispersed suspended solid we have a slug of suspended solid it will hurt the pump performance when this fluid gets into the centrifugal pump.

The creation of packet of gas/vapor or a slug of solid leads to making the flow non-homogenous, and when a non-homogenous fluid gets to centrifugal pump because

of non-homogeneity the pump starts to vibrate and in the long-term there will be failure in the system. The premature failure could be in the pump shaft, mechanical seals, etc. Therefore our main task is to make sure there is no creation of packet or a slug of gases/vapor or solids in the suction side of a centrifugal pump. Now we can discuss how harmful using a concentric reducer or eccentric reducer in the right position can be in the creation of a packet or slug of dispersed phase.

If there is a liquid-gas flow the best suction reducer is an eccentric reducer with the flat side of the reducer on top. To show it on a P&ID we show the eccentric symbol with a straight side on top and also we put an acronym of FOT representing "flat on top." If we are using a concentric reducer or an eccentric reducer with the flat on the bottom (FOB), in such cases the dispersed gas/vapor will accumulate in the tight corner of the reducer. This collected gas/vapor gradually gets larger and larger and eventually its size gets to a size that comes outside of the tight corner of the reducer. As soon as this happens the stream "sweeps away" the created pocket of gas/vapor and pushes it into the centrifugal pump casing. Also, when this happens the pump starts to heavily vibrate. Therefore, using the wrong type of suction reducer will cause frequent vibration of the centrifugal pump.

There would be the same problem if the wrong type of reducer is used for liquid-gas-solid two-phase flow. However, in this case as the suspended solid is heavier than the liquid they tend to accumulate on bottom of the pipe, and if the wrong type of reducer is used in the bottom tight corner of the reducer after a while a slug of solid gradually grows and finally sweeps away to our centrifugal pump casing and again heavy vibration happens. Therefore, if the service fluid is liquid-solid two-phase flow the best suction reducer is eccentric and FOB.

The P&ID representation of this discussion is shown in Figure 10.3.

Some people believe that the decision to use FOB or FOT depends of the elevation of the upstream reservoir connected to the pump. They believe if a pump gets suction from the bottom of a container, the reducer should

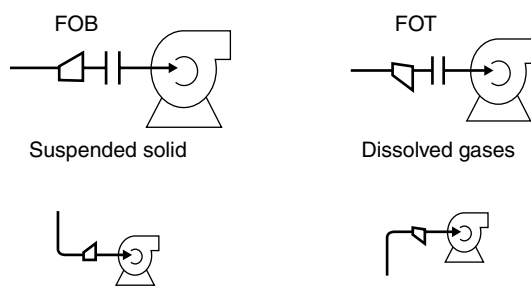


Figure 10.3 Acceptable eccentric enlargers on the suction side of pumps.

be FOB but if the pump gets suction from the top of a container, the reducer should be FOT. However the fundamental concept is what was mentioned above. When a pump gets suction from a bottom of a container, there is a chance of getting suspended solids and then we need an FOB reducer and similar logic for using an FOT.

When the suction flange of the centrifugal pump is smaller than the suction pipe size by more than one size, it is recommended to use multiple reducers in series instead of using one reducer to decrease the size to match with the suction flange of the centrifugal pump. The reason is that a reducer of reduced pipe size by more than one size may generate some disturbance in the liquid and this disturbed liquid, when it gets to the centrifugal pump, cannot be efficiently pumped. In such cases, two reducers in series but not back to back should be used.

Figure 10.4 shows a P&ID representation of a centrifugal pump with the associated reducer and enlarger.

There could be a strainer in the suction of a centrifugal pump. Installing a strainer in the suction of a centrifugal pump is very common if the installed pump is not supposed to receive large chunks of suspended solids in the liquid.

A strainer on the suction side of a centrifugal pump could be placed for a short term period or a long term period. If the strainer is used for a short term period it can be named as a TSS or “temporary suction strainer.” It means this strainer should be in place only temporarily during commissioning and then the start-up. Commissioning, which is the first start-up of a unit or plant after the construction, is different from other start-ups during the lifetime of a unit or plant. Because during the construction phase of a plant all vessels are open and pipes are open there could be the chance of a large chunk of solids in the system. These solids could be anything from used welding rods, instrument packages, or even

socks. Therefore during the commissioning a pump may see a large solid that could be detrimental for the pump internals, including impellers. So it is a very good idea to place a strainer on the suction side of a centrifugal pump temporarily during commissioning.

However, there are some cases that for whatever reason there is a still chance of having large solids in the pumping liquid. In such cases the strainer could be placed permanently and during normal operation of the pump.

The size of a strainer opening is decided based on the smallest clearance in the pump. It is obvious that some centrifugal pumps that are designed to handle large solids like slurry pumps, or some submersible pumps, don't need a strainer on their suction side.

10.6.1.2 P&ID Development on the Discharge Side

On the discharge side of a pump (the pump's downstream), there could be an enlarger and most likely a pressure gauge, a check valve, and also an isolation valve. After the isolation valve there could be a control loop to control the capacity of the pump. A Tee may exist for minimum flow spillback. The spillback is discussed in Section 10.6.2.

The check valve is a very critical component of a centrifugal pump and it should always be installed on the discharge side of a centrifugal pump.

The reason for requirement of a check valve is to prevent backward rotation of the impeller in the centrifugal pump when there is a sudden trip in the pump. When there is a sudden shutdown in the pump, the pump won't rotate and it will stop; however, the pumped fluid on the discharge side of the pump then will no longer be pushed and it may travel back from the discharge side of the pump and into the pump. When the discharge side of a centrifugal pump is a large pipe and/or it is a long pipe the severity of the backward rotation of the impeller is higher. In such cases it may be decided to insert a non-slam check valve. Backward rotation of the impeller in the pump is bad for at least for two main reasons: it makes the mechanical seal fail and also back rotation is bad for electric motor. A check valve should always be placed on the discharge side of a centrifugal pump and as close as possible to the discharge flange. The criticality of the distance between the discharge flange of a centrifugal pump to the check valve depends on the bore size of the discharge pipe; the larger bore size the more critical it is to keep the check valve closer to the discharge flange.

The last item in the centrifugal pump arrangement is isolation valves and blinds. Centrifugal pumps, isolation valves, and blinds should be used on both sides.

Up to now, a typical P&ID representation of a centrifugal pump could be like that shown in Figure 10.5.

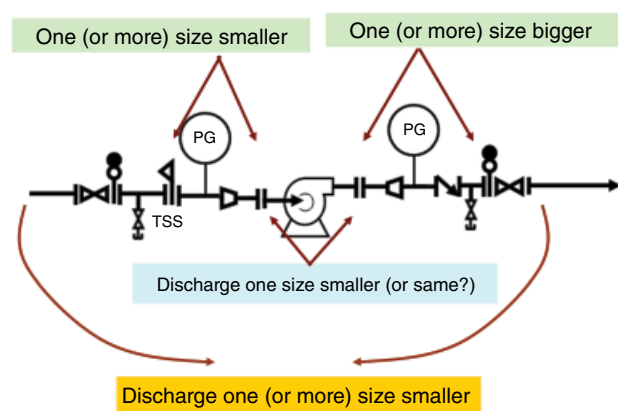


Figure 10.4 Centrifugal pump with associated reducer/enlarger.

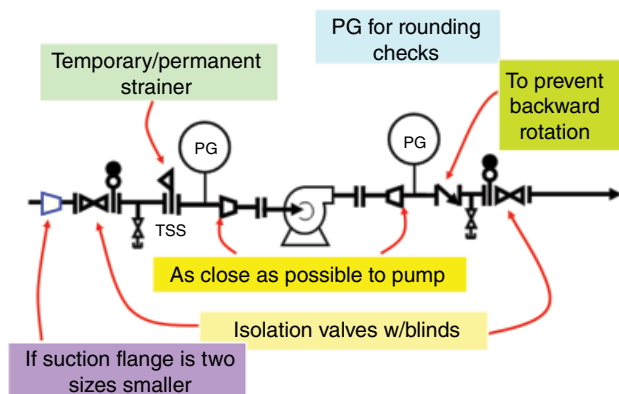


Figure 10.5 P&ID representation of a centrifugal pump.

To explain the rest of the requirements for developing a P&ID for centrifugal pumps, two main weaknesses need to be discussed.

Centrifugal pumps have two main problems: cavitation and low flow intolerance. When there is a weak point in a piece of equipment, this should be addressed through good process design and/or good control design.

When you are buying a centrifugal pump with a rated capacity of $400 \text{ m}^3 \text{ h}^{-1}$, the seller tells you: “Thank you for buying this and by the way, if you want to enjoy this new pump, be careful about two things in particular (amongst a bunch of other things!): first, this pump has a minimum flow rate of (for example) $100 \text{ m}^3 \text{ h}^{-1}$, so please don’t feed it with a lower flow rate than that.”

And second, be careful about the net positive suction head (NPSH); this pump has a “required NPSH” of (for example) 2 m, so please don’t feed it with a liquid with less than 2 m of “available NPSH.”

In Section 10.6.2 these two weaknesses will be discussed together with system process design and control design methods to mitigate these weaknesses.

10.6.2 Low Flow Intolerance and Minimum Flow Protection System

“Minimum flow rate” is a flow rate that is reported by the pump manufacturer, and if the pump receives a flow rate lower than that, it will start vibrating, heating up, and in the long term the pump will experience premature failure.

A pump’s minimum flow is provided by the manufacturer; however, as a first, rough estimate 30–40% of the “BEP flow rate” can be considered as the approximate minimum flow of a centrifugal pump.

During the operation of the plant the flow rate may go to values even lower than the minimum flow rate; how can we protect our pump when the flow rate goes below the minimum flow rate of our pump? To protect the pump from flows below the minimum flow reported by

the manufacturer, a specific arrangement including a minimum flow recirculation pipe (or “spillback”) with a control system needs to be implemented. This system can be called the “minimum flow protection” system.

If, for whatever reason, the flow to a pump is decreased to a value lower than the minimum flow rate of the pump we generally don’t have any control over increasing it if this reduced flow rate is inevitable. What we can do is basically a trick, and is the recirculation of a portion of the stream around the pump to “fool” the pump into “seeing” a flow rate higher than its minimum flow rate. It is important to know that by recirculation of flow around the centrifugal pump we just increase the flow rate in one circle and include the pump but this trick doesn’t increase the flow rate beyond the recirculation loop, it’s upstream or it’s downstream. This technique is named the “minimum flow protection pipe” or “minimum flow recirculation pipe” or “minimum flow spillback.”

Now that we have learned that by recirculation we can protect a pump against thermal and mechanical instability during very low flow rates, a few questions should be answered to be able to implement this trick.

The four questions that should be answered to be able to implement a spillback system on a centrifugal pump are:

- 1) In which pumps do we need to implement a recirculation loop?
- 2) Where should we position the recirculation pipe?
- 3) What should be the destination point of the recirculation pipe?
- 4) What should the size of recirculation pipe be?
- 5) What should the arrangement of the recirculation pipe be?

We are going to answer all these questions one by one.

10.6.2.1 Which Pumps May Need a Minimum Flow Pipe [1]

From what it was mentioned about the requirement of a minimum flow recirculation pipe one may believe that this is necessary for all pumps. In practice quite a few centrifugal pumps in a plant may have a recirculation pipe. However, there are some cases that may not need a minimum flow recirculation pipe. They are: pumps in closed circulating systems, pumps in intermittent services, small pumps, and pumps with a control valve on the flow loop or valuable speed device (VSD).

Pumps in fully closed circulating systems may not need a minimum flow pipe. The reason is that flow in completely closed circulating systems doesn’t change. In such systems there is no flow-in or flow-out and therefore flow is always constant. If flow is constant why should we put a minimum flow pipe to protect the pump against very low flows? Examples of such systems are hot water systems for HVAC purposes (heating, ventilation,

and air conditioning). There are several other examples of fully closed systems in the HVAC industry.

Pumps in intermittent services may not need a minimum flow pipe. It is not very common to see a pump in a plant that works in intermittent service while at a very low flow rate. Generally speaking the problem of low flow rate in such situations can be resolved by changing the schedule of operation for the intermittent pump. There could be a mistake in the design that an intermittent pump works at very low flow rate, lower than the minimum flow rate. In such cases, the design should be corrected.

We may not need to insert a minimum flow protection pipe for very small pumps. Small centrifugal pumps, say less than 5 hp may not need a spillback system. It is just for the purpose of lowering the required capital cost. The lack of implementation of a spillback protection system for small pumps is not because they don't see any low flow rate. They may see low flow rates and it is detrimental for them; however, losing them and the cost of replacing them is so low that it doesn't justify the implementation of a more expensive spillback protection system.

If a pump works with a flow control loop on its discharge side it may not need a spillback protection system. When a pump has a flow control loop the set point of the loop is definitely a value higher than its minimal flow rate. Therefore such pumps may not need a spillback protection system. This is the case for pumps with a VSD on them too. If there is a VSD on a pump the rpm of the electric motor is changed based on the current flow rate. When flow rate is low the rpm will be decreased by the VSD system and the good news is that the required minimum flow of the pump will be decreased accordingly. Therefore, we may not need a spillback protection system for centrifugal pumps with VSD control. However, there are some companies that don't like not implementing a spillback protection system based on the idea of a flow control loop or VSD system.

Based on the above discussion a spillback protection system may be required for other cases. These cases are pumps in mainstream and continuous operation, large pumps (larger than 5 hp), and pumps with level control loops.

10.6.2.2 Where Should we Position the Recirculation Line?

The second question that should be answered regarding the minimum flow protection pipe is about the takeoff point. There are primarily three points that a minimum flow pipe can start at; upstream of the check valve, between the check valve and the isolation valve, and downstream of the isolation valve. The second option, between the check valve and the isolation valve, has no process benefit therefore we focus on upstream of the check valve or downstream of the isolation valve (Figure 10.6).

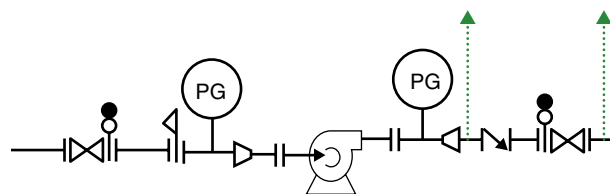


Figure 10.6 Two potential locations of the minimum flow pipe take-off point.

To satisfy the goal of a minimum flow protection pipe the takeoff point could be either upstream of the check valve or downstream of the isolation valve; both of them work well. However, if the takeoff point is placed upstream of the check valve it provides an additional benefit. If the isolation valve mistakenly gets closed the dead head condition (shut-off condition) happens. Generally speaking the dead head condition is not detrimental for centrifugal pumps or electric motors. It is not even detrimental for discharge piping if the design pressure of the piping is decided properly. However, if one centrifugal pumping system is kept under dead head condition for long period of time the trapped liquid will get gradually hotter and hotter. This is because of the addition energy of the trapped liquid through the pump's shaft. This increase in the temperature of trapped liquid may not be very harmful for non-flammable liquids; however, if the pumping liquid is a flammable liquid there could be the chance of fire. Because of that, a good idea is that the minimum flow protection pipe starts from upstream of the check valve rather than downstream of the isolation valve. Some companies prefer to follow the same rule for larger pumps even when the pumping fluid is not flammable. Their logic is that if a pump is large the trapped liquid during the dead head condition will quickly get hot and this is bad for the methodology of the pump casing.

10.6.2.3 Where Should the Destination Point of the Recirculation Pipe Be?

The next question about the minimum flow protection pipe is the destination point. The best destination point, which is consequently the most common one, is the upstream container. This means that best practice is to route the minimum flow protection pipe to the upstream tank or vessel. However, this is not always practical and/or economical. If the pump is very far from the upstream container and/or the minimum flow pipe is large enough, it is possibly not a good idea to route the minimum flow protection pipe to the upstream container.

The other option in such situations is routing the minimum flow pipe to the suction side of the pump. In such situations generally the minimum flow pipe is tied in to somewhere downstream of the suction isolation valve.

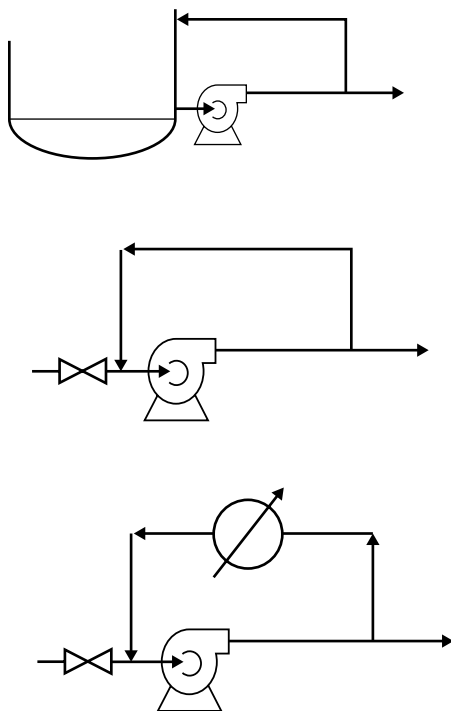


Figure 10.7 Spillback destinations.

The bad thing about this practice is the small recirculation of liquid around the pump absorbing the pump's power. This causes an increase in temperature of the trapped but recirculating liquid inside of the minimum flow pipe and the pump. This is an especially bad practice if the liquid is a flammable liquid. One solution for this issue is installing a cooler on the minimum flow protection pipe. However, this addition to the system makes this option less favorable in comparison to routing the minimum flow protection pipe to the upstream container.

Different alternatives of minimum flow pipe destinations are shown in Figure 10.7.

10.6.2.4 What Should the Size of the Recirculation Pipe Be?

The last question about the minimum flow protection line is pipe size and the control system on the pipe.

The minimum flow protection pipe should be sized based on the minimum flow of the pump, which is provided by the pump vendor. However, this number is not available until the contract has been awarded to the vendor and they have received their documents. This could be too late to size this pipe. To avoid delay in sizing pipes the minimum flow protection pipe is generally sized based on 30–50% of the pump design capacity. Here the sizing of this pipe is not very critical, the reason being that the power for moving the liquid is already provided by the pump, therefore if pipe is narrower than usual it is acceptable. Some companies allow the design to violate

their pipe sizing guidelines for sizing of the minimum flow protection pipe by using higher velocity in the pipe for the same reason.

As a rule of thumb the recirculation pipe for minimum flow protection could be half the size of the discharge pipe, or a size larger. For example, if discharge pipe size of a pump is 8", the minimum flow pipe could be 4" or 6".

10.6.2.5 What Should the Arrangement on the Recirculation Pipe Be?

Now it is important to consider how to implement the control system on the minimum flow protection pipe. Without a control system the majority of the liquid on the discharge side of the pump, when trying to find the least resistance route, will go through the minimum flow pipe rather than going toward the destination of the pump.

The most common type of control system on minimum flow protection pipe is a flow control loop (Figure 10.8). This loop includes a control valve on the minimum flow pipe, a controller with the set point of minimum flow provided by the pump vendor, and a flow sensor on the discharge side of the pump and upstream of the takeoff point. It is very important to consider that the flow sensor should be upstream of the takeoff point or basically inside the minimum flow protection loop and not outside it. It has been seen that in some designs the designer mistakenly puts the flow sensor downstream of the takeoff point or outside of the minimum flow protection loop. This design doesn't work properly because the flow sensor outside of the minimum flow protection loop senses the real flow rate, which is not compensated for minimum flow instability in the pump. Therefore, during the low flow conditions it keeps sending signals

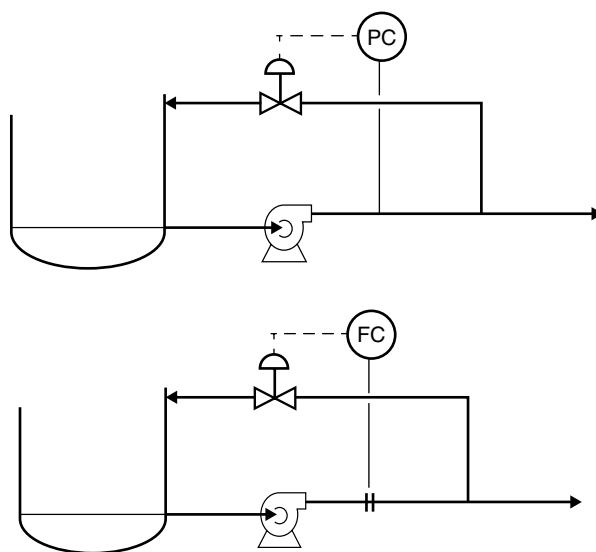


Figure 10.8 Minimum flow protection pipe with control loop.

to the control valve to open and open further, and this is why it doesn't keep up with the low flow.

The other option is a loop similar to the flow control loop but a pressure control loop. In this design, instead of a flow sensor we install a pressure sensor and the loop is a pressure loop. This control system doesn't work for all conditions as it only can work when the pump curve is adequately steep.

It is generally justifiable to use a control system on the minimum flow protection pipe for larger pumps, say larger than 35 hp.

For smaller pumps a similar system could be used, but instead of a control system we can have an on/off system. This means the flow sensor is still used but instead of a control valve a switching valve can be used. Whenever the pump flow goes below the minimum flow this valve will be opened to recirculate some flow to increase the pump flow rate to a value higher than the minimum flow. As you can see this is not a very accurate way to adjust the flow to protect the pump against the minimum flow conditions. However, for smaller pumps, say less than 35 down to about 20 hp, it is not justifiable to use a more expensive control loop and instead this switching loop is used (Figure 10.9).

For smaller pumps less than 20 hp a continuous minimum flow is implemented for the pump. In such cases just a restrictive orifice (RO) could be used on the minimum flow protection pipe (Figure 10.10).

For very small pumps, say less than 5 hp, as was mentioned there is no need to consider implementing a minimum flow pipe at all.

It is important to realize that a minimum flow protection pipe is not working during the majority of pump operation. It is because in the majority of times the control valve on the minimum flow protection pipe is closed. This valve will be open only when the flow to the pump

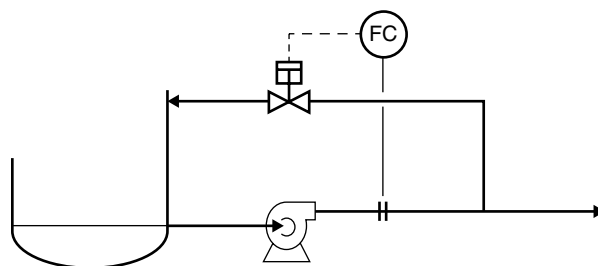


Figure 10.9 Minimum flow protection pipe with switching loop.

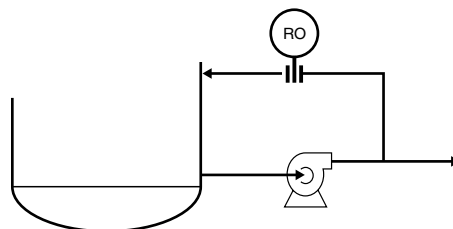
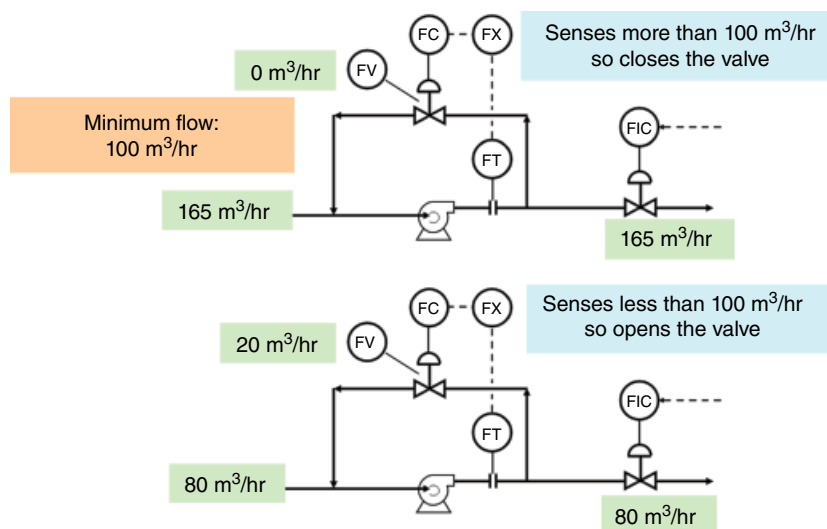


Figure 10.10 Minimum flow protection pipe with restrictive orifice.

goes below the minimum flow value (Figure 10.11). This is the reason that some companies violate their pipe sizing criteria and use higher velocity design bases for a pipe to come up with a narrower and cheaper pipe for this purpose.

However, the control valve on minimum flow protection pipe should be “failed-open” (FO). This valve should be FO because its function is important to keep the pump operational. When losing the instrument air if this valve is “failed-closed” (FC) the pump will start to vibrate if the flow goes below the minimum flow value. So to keep the pump operational this valve should be FO.

Figure 10.11 Functioning of a minimum flow protection pipe.



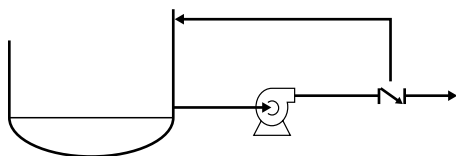


Figure 10.12 Minimum flow protection pipe with an automatic recirculation valve.

All the systems on the minimum flow protection pipe could be replaced with a device named an automatic recirculation valve or ARC. It looks very attempting to use a single device such as ARC instead of the more complicated, utility dependent control system. However, ARCs have some shortcomings. First of all they are very prone to plugging if the service liquid is not clean. The other limitation is that ARCs are not available in large sizes, possibly larger than 6". Inside ARCs are a bypass valve and a spring-loaded check valve. The P&ID symbol for an ARC is very similar to a check valve but the difference is that the ARC symbol has two outgoing lines (Figure 10.12).

It is important to know that not all loops around the pumps in the P&ID are "minimum flow protection pipes," there could be similar pipes in the P&ID or plants for some other reasons. A loop around the pump could be implemented on a pump for the following reasons:

- As a minimum flow protection pipe.
- To protect the pump from gradual heating during dead head conditions (mainly for flammable liquids).
- To provide fluid moving around the pump to maintain the flow during electrical outages where the pump is supplied by emergency electricity.
- To achieve an acceptable efficiency in very small centrifugal pumps (less than $10 \text{ m}^3 \text{ h}^{-1}$).
- To providing a start-up pipe for positive displacement pumps.

If this loop around a pump is implemented for each of above reasons, they should be sized based on different criteria.

10.6.3 Cavitation

Cavitation is a phenomenon that is related to the generation and collapse of vapor bubbles inside of pumps.

When this happens, the pump fails prematurely because of bubbles slamming into the impeller and the internal side of the casing. The collective name for these events is cavitation.

The main underlying reason for cavitation is a lack of enough pressure on the suction side of a centrifugal pump. When a centrifugal pump operates, it basically "sucks" the liquid, which can generate bubbles on its suction side, causing cavitation.

Have you tried to suck a carbonated drink out of a tall bottle with a narrow straw? You might notice a bunch of bubbles coming into your mouth. However, luckily cavitation won't happen in your mouth because bubbles don't have much speed!

If the liquid on the suction side of a centrifugal pump has "enough" pressure, it won't release gas and cavitation won't happen. The problem of a "lack of enough pressure on the suction side of a centrifugal pump" can be stated more technically as: a "lack of enough NPSH."

NPSH or "net positive suction head" is basically the total "effective" pressure of a liquid at the suction flange of a centrifugal pump in "head" units (e.g. meters or feet).

Each centrifugal pump has a minimum acceptable NPSH, which is reported by the pump manufacturer and is termed the "required NPSH," or net positive suction head required (NPSH_R).

A typical NPSH_R for a centrifugal pump could be a value anywhere from less than a meter, up to more than 10 m.

After buying a centrifugal pump, the process and the control system should be designed in order to ensure that the liquid has enough pressure at the suction flange to prevent cavitation. This pressure, which is provided by the system (rather than the pump), is termed "available NPSH," or net positive suction head available (NPSH_A), and is reported in the same units as head (e.g. meters or feet).

This concept is shown in Table 10.7.

NPSH_A should be higher than NPSH_R by a pre-selected margin, otherwise the centrifugal pump will most likely cavitate.

The service fluid type is part of the "system" too. Where the pumping fluid is hotter or more volatile the pump is more prone to cavitation.

There are, however, cases where NPSH_A is lower than " $\text{NPSH}_R + \text{margin}$." There different techniques available to solve the problem. To apply some of the solutions the designer should go back to the design stage of project but for some others, some changes during the P&ID development could solve the problem.

Each of the items in Table 10.8 need detailed evaluation by process engineers and other stakeholders to check their applicability.

A "stand pipe," which is a solution to increase NPSH_A , is a simple vertical vessel that accumulates the liquid to a higher level for the benefit of the downstream centrifugal pump. The dimension of a standpipe is decided and these rules of thumb can be used. The diameter is preferable less than 24" to be able to use a piece of pipe (seamless) as the body of the standpipe. The height is primarily defined by the required level of liquid in it to provide enough NPSH_A . It is generally preferable to leave the top of the standpipe open to atmosphere to get rid of

Table 10.7 Relationship between $NPSH_A$ and $NPSH_R$.

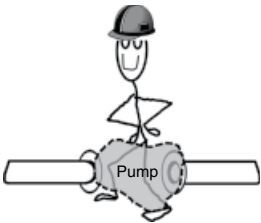
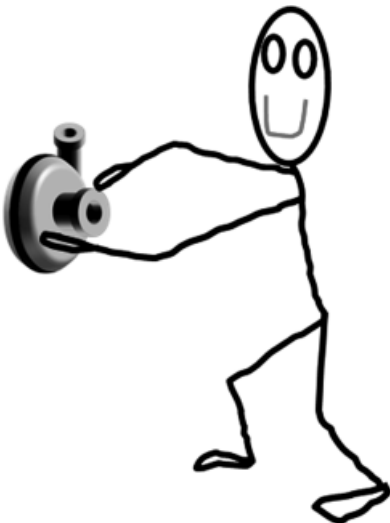
Parameter	$NPSH_A$	$NPSH_R$	Comment
Provided by	 		N/A
	Plant/System	Pump manufacturer	
Example 1	3 m	> 2 m	Ok, you are good to go! Your $NPSH_R$ is lower than your $NPSH_A$! You are lucky!
Example 2	1.2 m	< 3 m	Oops! Your $NPSH_A$ is lower than your $NPSH_R$! You need to do something: 1) Ask your process engineer to somehow increase the $NPSH_A$ 2) Or shop around to find another less restrictive pump with a higher $NPSH_R$

Table 10.8 Solution for the issue of $NPSH_A$ being too close to $NPSH_R$.

How to increase $NPSH_A$ Changing system features	How to decrease $NPSH_R$ Changing pump features
<ul style="list-style-type: none"> ● Raise the liquid level (e.g. using a stand pipe) ● Raise the source container elevation ● Lower the pump and put in a pit ● Decrease the suction pipe length ● Increase the diameter of the suction pipe ● Eliminate fittings in the suction pipe as much as possible ● Increase the gas pressure in the source container ● Use two pumps in series (booster) ● Decrease the liquid temperature (e.g. by injecting cold liquid) 	<ul style="list-style-type: none"> ● Use a low rpm pump ● Use double suction impellor ● Use an oversize pump ● Use two smaller pumps in parallel

complexities related to the need for pressure safety valves (PSV). If this is the case, the height of the standpipe should be higher to work as a “cold vent” too. If an open top standpipe is not possible, other options should be incorporated to make sure the design is safe.

After finishing the main topics on the development of centrifugal pumps we deepen our understanding by learning other concepts related to centrifugal pumps in Sections 10.6.4–10.6.11.

10.6.4 Very Small Centrifugal Pumps

When the required capacity of centrifugal pumps is small (say less than $10 \text{ m}^3 \text{ h}^{-1}$), their efficiency decreases. This is especially true for centrifugal pumps with capacity of less than $5 \text{ m}^3 \text{ h}^{-1}$. The low efficiency of such pumps is not only bad because they are wasting energy, the pumps also wear out more quickly. Therefore provision should be considered to prevent that.

One solution is to buy a larger capacity pump and then provide a continuous recirculation stream around the

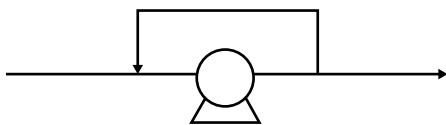


Figure 10.13 Relationship between $NPSH_A$ and $NPSH_R$.

small centrifugal pump (Figure 10.13). This can be done by a pump that connects the discharge to the suction side of the pump. The flow in the pipe can be limited by an RO (restrictive orifice) on the route. This solution obviously doesn't prevent waste of energy but it helps the pump to last longer.

When implementing this solution, it is necessary to check the ultimate temperature of the liquid inside the pump and also toward the pump downstream. It should be ensured that the temperature is still low enough that it cannot support liquid boiling and also the hotter liquid is acceptable for the downstream equipment. If the temperature increase is not marginal and is not acceptable, the recirculating pipe may be directed to the upstream container rather than the pump suction side.

10.6.5 Different Types of Spare Pump

Although the discussion in this section is about pumps, the rules apply to any other spare items.

It is very common to see that pumps, mainly in the mainstream, have spares.

When spare pump(s) is (are) available, the question is whether each single pump can work as an operating pump and when it is needed as spare pump or not. Although this is the best option, there could be cases where a pump only serves in the system as a spare pump and shouldn't be in operation for extended periods of time. In the second option there is a "designated spare pump" while in the first option each single pump functions as a "non-designated pump" or operating/spare pump.

One common example in our daily lives is spare tires in our cars. The spare tires used to be exactly similar to "operating" tires and they could be used as an operating tire for extended times operating on a car when a tire was out of operation because of, for example, tire puncture. These days spare tires in cars are generally thinner and weaker than the operating tires. They cannot be operated for long periods of time. Therefore the spare tires in older days were the non-assigned type but nowadays are assigned types.

For spare pump(s) usually a client expects that all pumps should be able to act as an operating pump and a spare pump (non-designated spare pumps). However, in some designs, especially in the case of common spare pumps, providing a non-designated pump needs a more elaborate piping arrangement around the pumps. Generally speaking if a spare pump is depicted on a P&ID

it is assumed to be of non-designated type. If this is not the intention of the designer, a note should be added on the P&ID stating that the pump is a "designated spare pump."

From an asset management point of view it is not a good idea to leave a pump off for long periods of time. An assigned pump that is left non-operational for a long time couldn't be brought into operation with no problems and as quick as possible.

There are some cases that there are some pumps working and one or more pumps as "workshop spare(s)." In this scenario the designer/client does not provide a spare fully available but they store one pump ready to be installed in the case of losing the operating pump(s) within a short time (e.g. 24 hours). Such an arrangement cannot be shown simply and can only be captured by a note in the call-out of the pumps. A spare pump, depending on the criticality of the service, can be an "installed spare" or a "workshop spare."

A workshop spare cannot be seen on a P&ID other than a note beside the main pump stating the existence of workshop spare.

When one pump serves as a spare pump for two or more pumps (in designated or non-designated duties) their sparing functionality could be in the form of "dedicated spare" or "non-dedicated spare" type. A dedicated spare pump can work as spare if only one specific pump goes out of service while a non-dedicated spare pump can be brought into service when any of the parallel pumps goes out of service. A dedicated spare pump, basically, provides a spare for only one other pump but a non-dedicated (or common) spare pump provides a spare for more than one pump.

In all the above cases a spare pump could have the same capacity as the operating pumps or less than them. In such cases, when there is a need to use the spare pump, the flow of the system is dropped. This type of spare could be named a "partial spare." This could be the case for very expensive cases where full capacity pumps cannot be afforded.

The decision for any of above options is based on different parameters including:

- Mean time between failure (MTBF) of the equipment
- Mean time to repair (MTTR) of the equipment
- Cost of maintenance
- Value of the "lost production."

This is generally the decision made by the asset management mechanical engineer and may come from a RAM (Reliability, Availability and Maintainability) analysis.

10.6.6 Centrifugal Pump Arrangements

There are two arrangements for centrifugal pumps: centrifugal pumps in parallel and centrifugal pumps in series.

10.6.6.1 Centrifugal Pumps in Parallel

For parallel arrangement of pumps there are different arrangements including the operating-spare arrangement, the lead-lag arrangement, and simultaneously operating pumps.

Centrifugal pumps can be arranged in parallel forms for different reasons. They are:

- If the flow rate that should be handled by the pump is huge and there is no available single pump to be used, two or more centrifugal pump can be placed in parallel instead of one single pump.
- When a higher reliability is needed: this is the case that we need to put a spare pump in parallel with the main pump. However, in this case there is no time that all parallel pumps work together.
- When enough $NPSH_A$ is not available: sometimes the suction side of a centrifugal pump doesn't provide enough $NPSH_A$ and it is close to or less than the $NPSH_R$ one solution is using two or more smaller centrifugal pumps. This can solve the problem because smaller pumps have smaller $NPSH_R$.

Figure 10.14 shows a schematic of pumps in parallel. However, the designer should be careful about recognizing centrifugal pumps in parallel. They are not always easy to recognize.

If there is a single pump pumping a liquid from point A to destination B and somewhere in the middle of discharge pipe the discharge of another centrifugal pump is merged to the pipe, these two pipes are considered as parallel pumps. The importance of recognizing parallel pumps is that they usually are through one single datasheet. When two centrifugal pumps operating as parallel pumps they should be completely identical. Because of that they are usually placed in one single datasheet to make sure they are from one single pump vendor and they are identical.

10.6.6.1.1 Minimum Flow Pipe for Parallel Pumps

If there are parallel pumps but only one of them works at the time and the rest are spare pumps there is no concern about the minimum flow protection pipe. In this case, the minimum flow protection pipe can be branched-off from the main discharge header of the pumps. However, when there are operating parallel pumps the concept is a

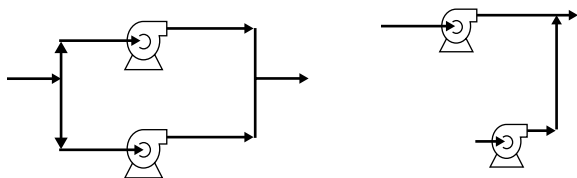


Figure 10.14 Parallel pumps.

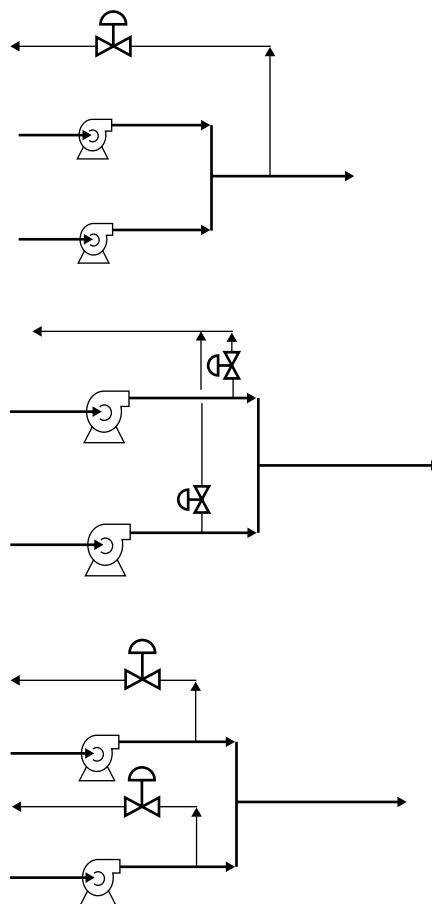


Figure 10.15 Minimum flow pipe for parallel operating pumps.

little bit more complicated. There are at least three available options (Figure 10.15).

Option 1: one minimum flow protection pipe as a shared minimum flow protection pipe for all the parallel operating pumps. Although this option is acceptable when only one parallel pump is operating at a time it is not a good practice for simultaneously parallel operating pumps. The reason that this is not a good idea is that this option doesn't necessarily compensate the starving pump. As the parallel operating pumps don't have dedicated minimum flow protection pipes if one of the pumps operating with the flow rate below the minimum flow rate and the other pump operating well, this system cannot work properly. This option could be improved with a good control system. The control system for this option will be discussed in Chapter 14.

Option 2: Dedicated minimum flow protection pipe for each parallel pump but with a merged pipe after the control valve. In this option the better approach of using a dedicated pipe and a dedicated control valve is used; however, to save some money all the dedicated

minimum flow pipes will be merged together and as a single pipe route to the destination. This option is a technically and economically acceptable option and is possibly the most common type of arrangement for a minimum flow protection system of parallel operating pumps.

Option 3: dedicated minimum flow protection system including pipe and control system for every single operating parallel pump. This arrangement is the most expensive arrangement and should be used only if the service is very critical. In this option the system of parallel pumps are seen as their elements. No cost saving is implemented and a minimum flow protection pipe is implemented for each of the pumps totally independently.

In the case that each pump has own dedicated minimum flow protection pipe there could be a chance of reverse flow. The reverse flow could be from the operating pump back to the discharge side of the idle pump and then going to the pump in the reverse direction. So provisions should be implemented to prevent this. The provision is basically installing one check valve for each minimum flow protection pipe. However, the location of the check valve can be decided. There are two alternative arrangements available.

Alternative A: in this alternative one more check valve should be installed upstream of each minimum flow branch of point. This should be done for all parallel pumps. In this alternative there are basically two check valves on the discharge side of each centrifugal pump; one upstream of the minimum flow branch of point and the other one downstream of the minimum flow branch of point (Figure 10.16).

Alternative B: in alternative B check valves are installed on each single minimum flow protection pipe to prevent backflow. This option is a less expensive option as

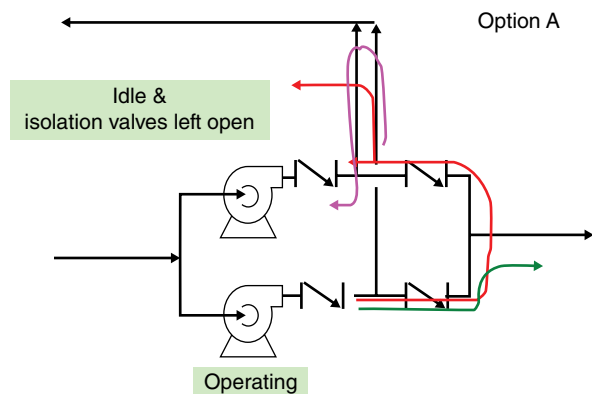


Figure 10.16 Preventing reverse flow through minimum flow pipe of parallel pumps (alternative A).

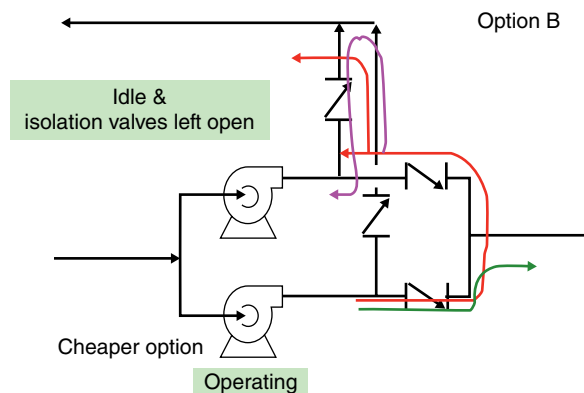


Figure 10.17 Preventing reverse flow through minimum flow pipe of parallel pumps (alternative B).

the check valves should be installed on small bore minimum flow protection pipes rather than large bore discharge pipe in alternative 1 (Figure 10.17).

10.6.6.2 Centrifugal Pumps in Series

Sometimes centrifugal pumps should be placed in series with each other.

It is important to recognize that centrifugal pumps in series are not always easily noticed. One common arrangement of two centrifugal pumps in series is the case where a second pump takes suction from the discharge of the first pump. In that case, these two pumps also are considered as centrifugal pumps in series (Figure 10.18).

The second reason for using centrifugal pumps in a series arrangement is to provide a very high discharge pressure. Generally speaking a centrifugal pump can generate a discharge pressure of more than 3000 kPag. If a higher pressure is needed centrifugal pumps can be installed in a series arrangement. However, putting more than two centrifugal pumps in a series arrangement is

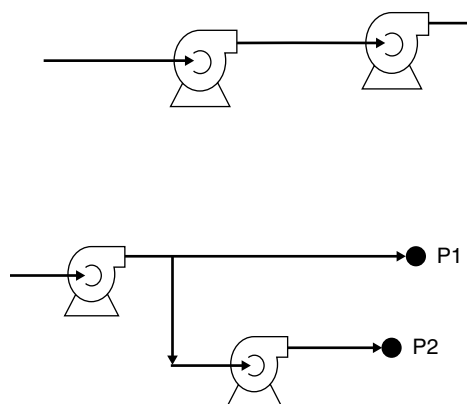


Figure 10.18 Pump in series.

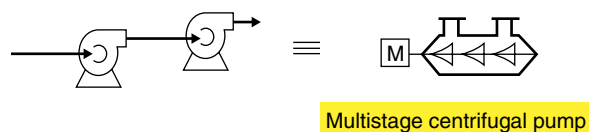


Figure 10.19 Centrifugal pumps in series and multi-stage centrifugal pump symbols.

not common. If there is a need to install three or more centrifugal pumps in series to attain to a specific discharge pressure, a better option is using a single multi-stage centrifugal pump. A multi-stage centrifugal pump is equivalent to multiple centrifugal pumps in series. In a multi-stage centrifugal pump each stage represents one single centrifugal pump (Figure 10.19).

One common example of using pumps in series is when a pump system doesn't have enough $NPSH_A$. One solution for a shortage of $NPSH_A$ is to put another pump upstream of the main pump to pressurize the liquid to provide adequate $NPSH_A$ for the second pump. In this arrangement the pump with the duty of providing $NPSH_A$ is named a booster pump and the second pump can be named things like main pump or transfer pump. Generally speaking a booster pump should have the same capacity (or flow rate) of the main pipe but its discharge pressure is much lower than that. The discharge pressure of a booster pump could be four to seven times that of the required $NPSH_R$ or 350 kPag. One famous example is "boiler feed water system (BFW) pumps." BFW pumps provide high pressure water for the boiler. As water is of generally high temperature, the system doesn't have enough $NPSH_A$, therefore an arrangement of booster pump and main pump is needed for this purpose. From the other side, as high pressure is needed to the boiler, two pumps could be needed. Therefore a two pump arrangement of pumps in series with the names of "low pressure BFW pump" and "high pressure BFW pump" may be needed. The low pressure BFW pump works as a booster pump and a partial pressurizing pump. Therefore, the discharge pressure of low pressure BFW pump could be much higher than 350 kPag.

Here two issues regarding pumps in series are discussed: the minimum flow protection pipe and the check valve.

From a theoretical point of view of pumps in series only one minimum flow protection pipe is enough. The reason is that if one pump is suffering from low flow it will be the case for the second pump too (Figure 10.20). Therefore there is no need for the dedicated minimum flow protection pipe for each of the pumps in the series configuration. However, some companies are more conservative and like to see dedicated minimum flow protection pipes for each single pump in series of duration.

The other point about pumps in series is if we really need to put a check valve on the discharge side of each of the pumps in the series configuration. Again from

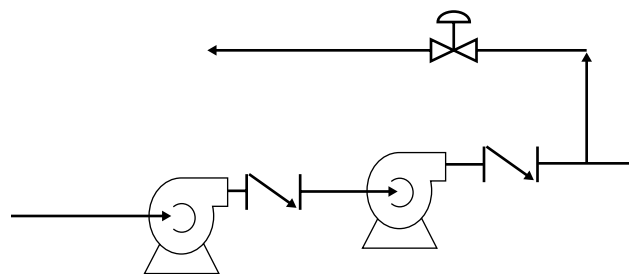


Figure 10.20 A sample P&ID for centrifugal pumps in series.

theoretical point of view one check valve is enough on the discharge side of the downstream pump. However, if the distance between the first pump and the second pump is large enough, another check valve for the upstream centrifugal pump may be needed. Again, some companies like to take a conservative approach and put check valve for each discharge site of pumps irrelevant of their distance.

10.6.7 Pump Warm-up or Cool-down System

For installed spares, if the ambient temperature of the space around the pump is far from the operating temperature of the pump, the pump should be a "hot stand-by" (or a cold stand by in cryogenic systems) pump to make sure the pump will not experience thermal shock during start-up.

Suppose there are two centrifugal pumps in parallel and one of them is an operating pump and the other one is a spare pump. The ambient temperature is very low and much lower than the pumping liquid temperature. There could be a case where operating pumps work for several days or weeks and at the same time the spare pump stays idle without any liquid in it. In such case all the components of the idle centrifugal pumps including impeller, casing, etc. will stay at the ambient temperature, which is very low. Now consider the case where the operating pump is suddenly tripped and doesn't work for some reason. In that case the spare pump should be placed in operation by an operator as soon as possible. However, by initiating the first amount of warm or hot liquid to the cold previously idle pump the pump will experience a very severe thermal shock. The thermal shock is because the wide difference between ambient temperature and the service liquid will cause cracks in the pump casing and the other components of the pump. Therefore, the previously idle pump could be out of work very quickly. How can we prevent such a bad failure?

There is a practice in process plants that when the difference between ambient temperature and the service liquid temperature is high (say more than 100–250 °C) a warm-up system should be considered to prevent thermal shock in the pump. This warm-up arrangement makes an idle pump ready for start-up without chance of quick failure of the pump.

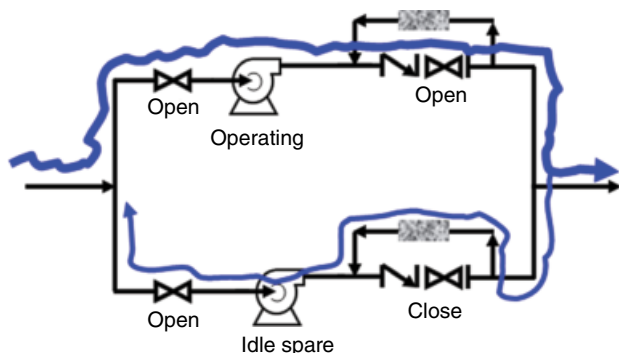


Figure 10.21 Warm up arrangement 1.

There are at least two arrangements to implement the concept of a pump system. The one which is the more popular arrangement is shown in Figure 10.21.

In this arrangement a pipe loop bypasses the check valve and isolation valve in the discharge site of each pump in parallel if a spare pump is available. What this bypass pipe can do is allow a portion of liquid from the discharge part of the operating pump to be redirected and go slowly to the idle spare pump from its discharge site to its suction site. There are two important issues regarding this arrangement: first of all the isolation valve in the suction idle spare pump should be left open and the second one is that the bypass pipe should be narrow enough. The reason for installing a narrow bypass pipe is to make sure the flow that goes in the reverse direction in the idle spare pump is so low that it cannot back rotate the impellor of the idle spare pump. Otherwise the pump will be faced with some mechanical problems.

The way that we can make sure the flow to the idle spare pump is low enough is not only installing a small bore pipe but also putting ROs and/or a three quarter inch needle valve to adjust the flow.

In some operating plants where a pump warm-up bypass line was not inserted during the design they may solve the problem by the “duct tape” solution. This not very ideal solution is to drill some holes in the flapper of the check valve. The hold flapper of check valve works in a similar way to the warm-up bypass line. The reason that this solution is not very good is because the drilling generates some residual stress in the check valve flapper that may shorten the life of the check valve.

In some designs when we deal with very high viscous liquids an additional bypass pipe can be implemented in addition to bypass around the isolation and check valve. This bypass is from the discharge site of the centrifugal pump to the suction site of the centrifugal pump with a small needle valve on this or the RO or both of them (Figure 10.22).

The second arrangement to provide a pump warm-up circuit doesn't need anything to be implemented during

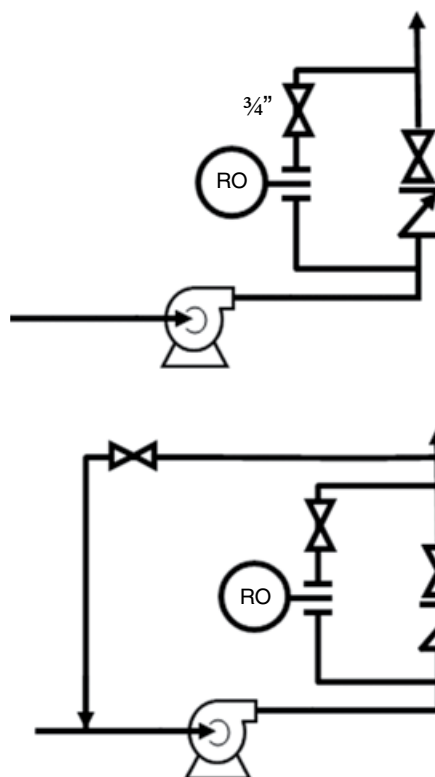


Figure 10.22 Variation of warm up arrangement 1.

the design. Therefore, it could be a good solution if no provisions are considered for pump warming-up during the design phase of the project. In this solution no additional hardware in the system is required. It only means that the suction isolation valve of the idle spare pump is left open together with the casing drain valve of the idle spare pump. When an operator does that a small fraction of flow is separated from the main flow to the operating pump and goes through the idle spare pump and exits from its drain valve. The bad thing about this warm-up arrangement is the loss of liquid from the system. In this solution a portion of service liquid is wasted through the drain valve of the idle spare pump. However, some people don't consider this as “real waste” as this wasted liquid eventually goes to the sump system and from the sump vessel it is returned back to the process system (Figure 10.23).

There is the same issue when a set of operating/spare pumps is working to pump a cryogenic liquid and the ambient temperature is much higher than the liquid temperature. This same arrangement should be considered to prevent the same problem of thermal chuck. However, in this case the arrangement is named a cool-down pipe rather than a warm-up pipe. Because the purpose of this arrangement will be keeping the idle spare pump at very low temperature, near the temperature of cryogenic liquid, to make sure when it is brought into the

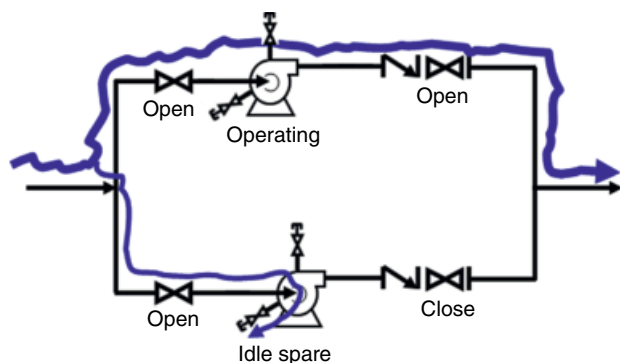


Figure 10.23 Warm up arrangement 2.

operation suddenly the pump doesn't crack because of thermal shock.

10.6.8 Piping Spec. for Centrifugal Pumps

Depending on the generated head by the centrifugal pump the pressure rating of the discharge flange of a centrifugal pump could be higher than the pressure rating of the suction side and suction flange. This is not the case for all centrifugal pumps but if the head of the centrifugal pump is very high the pump manufacturer may put a higher pressure rating for the discharge site. If this is the case a designer may decide to choose all the elements in the discharge side of the centrifugal pump with a pressure rating of the discharge flange. However, some companies that have a more conservative approach prefer to extend back the higher pressure rating of the discharge side of the centrifugal pumps up to the suction isolation valve of the centrifugal pump. This decision could be an especially good decision if there are multiple operating pumps in parallel. In that case there could be a chance that one pump is operating and pressurizes the suction side of the other pump (Figure 10.24).

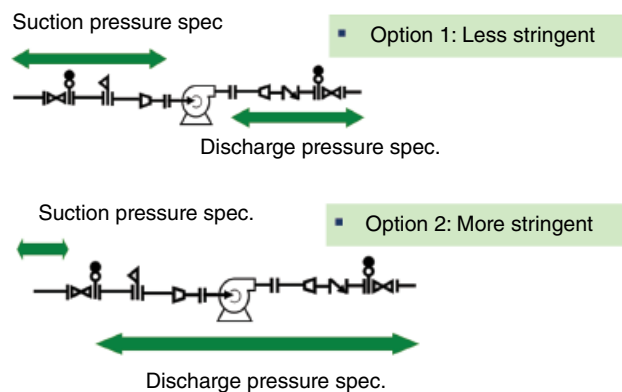


Figure 10.24 Extension of high design pressure class to the suction side of the centrifugal pump.

- Electromotor M
- Steam turbine (for hp > 100?) T
- Engine D G

Figure 10.25 Centrifugal pump drives.

10.6.9 Centrifugal Pump Drives

The most common type of drives for centrifugal pumps is electric motors. However, there are some cases that some other types of drives could be used, for example, a diesel engine or a gas engine could be used as a drive for a centrifugal pump. This is the case mainly for centrifugal pumps in remote areas where electricity may not be available. The other option is a steam turbine as a drive for a centrifugal pump. It is not very common to see steam turbines as drives for centrifugal pumps. However, it is more attractive to use a steam turbine as a drive for a centrifugal pump if the centrifugal pump is high power, say more than few thousand horse power.

Companies may or may not decide to show the pump drives. Figure 10.25 shows the P&ID representation of different drives.

However any "process drives" like turbines or motors need to be shown on the P&ID. It may be decided to show a diesel motor or a gas motor in auxiliary drawings rather than on the main process drawings.

10.6.10 (Liquid) Seal Systems in Centrifugal Pumps

Seal flush plans are supporting systems for mechanical seals in fluid movers (pumps, compressors, blowers, and fans). However, the question is: what is a mechanical seal?

In each centrifugal or rotary fluid mover there is a rotating shaft that is connected to the electric motor shaft (or the shaft of any other drive) from one side and from the other side it is connected to the fluid mover internals (for example the impellor). Therefore, the fluid mover shaft penetrates through the fluid mover casing. The penetration is done through a hole with a suitable diameter to provide enough room for the shaft to pass through the fluid mover casing and then to be connected to the driver.

Now the question is how loose or how tight this hole should be relative to the shaft diameter. If the hole diameter is much larger than the shaft diameter (loose system), there is always some leakage from this gap (Figure 10.26). If the diameter of the hole almost the same as the diameter of the shaft (tight system), there would be no leakage because there is no gap, but it is very difficult for the shaft to rotate in such a tight hole.

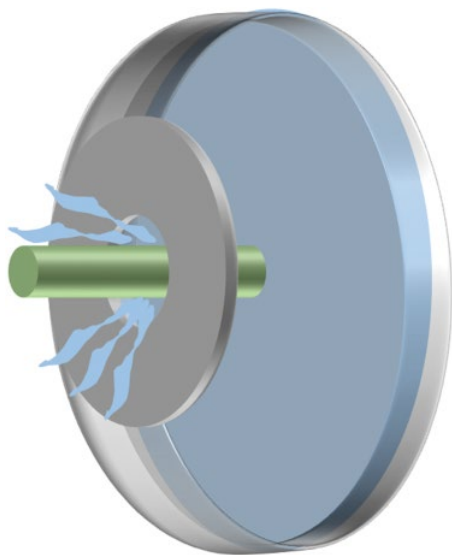


Figure 10.26 Loose shaft casing touch.

How can we resolve this problem?

The problem used to be resolved by application of a stuffing box and gland. In that solution, a small space is provided around the hole and donut shaped “packing” is placed inside of that. The packing in the first generations of stuffing box was a ribbon of felt. Therefore, felt from one side prevented the leakage from the other side; it provided enough freedom for the shaft to rotate.

The problem with a stuffing box and packing is that there is usually some leakage from the system. The other problem is that it needs frequent checking and inspection to make sure the packing is healthy.

This solution is still available for some pumps where the rotation speed of the shaft is not very high or the service is not very critical. It is interesting to know that valve stems are still handled by stuffing box and packing.

The second solution is a “mechanical seal” which is a more robust system. From a very simplistic viewpoint, a mechanical seal is two blocks of graphite sliding on each other. Both of these blocks could be in the form of donuts. One donut is fixed in the pump casing and around the hole and the second block of graphite is attached to the shaft at the penetration point. During operation of a fluid mover, these two blocks are sliding on each other. The features of the graphite blocks are: (i) they have very smooth surfaces and (ii) they inherently have some lubricity features because of the nature of graphite. Therefore, on the one hand, because of the very low roughness of the graphite blocks surfaces, there is almost no leakage from the system. On the other hand, because of the oily feature of graphite blocks, the shaft can easily rotate in the hole.

In reality, a mechanical seal is more complicated than the above and has several other elements. The two sliding

blocks plus other mechanical parts around them is named a mechanical seal. These days, there are more complicated mechanical seals that are constructed of different materials and sometimes graphite blocks are replaced by ceramic blocks, however, the fundamental operation remains the same.

Mechanical seals are a very critical part of every fluid mover and need frequent inspection.

However, a mechanical seal is not something that can be seen on a P&ID so why should we care about that in P&ID development? Even though there is no footprint of a mechanical seal on a P&ID each mechanical seal should have a supporting system and the supporting system should be shown on the P&ID. The supporting system for a mechanical seal is a system to flush, cool down the sliding blocks, and also lubricate those parts. Basically, the problems associated with mechanical seals are because there are sliding blocks, there is heat generation in sliding block surfaces, and this generated heat should be dissipated. By providing some sort of lubrication, the generated heat can be decreased. The supporting system that provides sealing, cooling, and lubrication for a mechanical seal is named a “seal flush plan.” There are plenty of different arrangements to provide cooling, flushing, and lubrication for mechanical seals.

The concept of sealing systems is applicable for all fluid movers. However, here we continue our discussion only for pumps. More discussion on sealing systems of gas movers is in Section 10.8.

There are different arrangements for sealing systems of pumps, which are standardized for the sake of decreasing the price and also facilitating fabrication and inspection. There is a standard generated by the American Petroleum Institute under the name of API-682 that introduced more than 30 different plans as supporting systems for mechanical seals. Usually, the mechanical engineer in each project decides on the selected type of seal flush plan based on standard plans stated in standards or a combination of a few of them, or even totally non-standard arrangements. Deciding on a specific mechanical seal flush plan is done through the flow chart that is provided in the standard. In standard API-682 each seal flush plan is specified by a number. Therefore, the seal flush plan could be introduced by seal flush plan number 13 or seal flush plan number 15 by referring to that specific standard.

Each seal flush plan needs a stream of fluid for the purpose of cooling, flushing, and lubrication. This flow stream can be provided from the process fluid, “the fluid that is being pumped,” or from an external source. The external source fluid can be plant water or some other type of fluid. The final destination of each flushing/cooling/lubricating fluid could be the process fluid or other destinations. To make sure that we have enough flow of

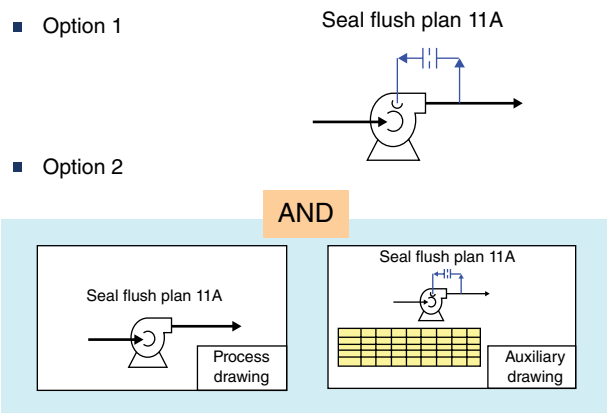


Figure 10.27 Two ways of showing a seal flush plan on P&IDs.

flushing/cooling/lubricating fluid to the mechanical seal, there is a bunch of sensors, manual valves, etc. on these arrangements. The instruments on seal flush plan should be tagged so that it can be captured by instrument engineers. Some seal flush plans need an external utility. The external utility could be “plant water” or other utilities. The P&ID developer should make sure the required utility for the seal flush plan is available. Otherwise they may need to change the seal flush plan if possible.

Now the question is if we need to show seal flush plan of each pump on each sheet of P&IDs or not. There are at least two ways to handle the presentation of a seal flush plan on P&IDs (Figure 10.27).

In option 1 all the arrangements related to the seal flush plan are shown on each single pump on a P&ID.

In option 2, we just show a reference number beside each pump that needs a seal flush plan and then we show the seal flush plan arrangement in an auxiliary P&ID drawing and tag all the instrumentation in the auxiliary drawing. The second option is the better option when the P&ID is crowded and showing each seal flush plan makes it more crowded and possibly non-usable. In the second option, we may just show the number of the seal flush plan number, for example the seal flush plan number 21 or 13, to refer the P&ID reader to the auxiliary drawing and to the seal flush plan 21 or 13. The second option has the advantage that all the pumps with the same seal flush plan can be returned to one single auxiliary P&ID that shows that specific plan. In that auxiliary P&ID, the arrangement can be shown and also a table below the arrangement shows the tag number for the element on the seal flush plan arrangement for each single pump or compressor. This way, we can leave a bunch of room on the P&IDs.

It is important to know that not all equipment with a rotating shaft needs mechanical seal and consequently a seal flush plan system. In some low RPM positive displacement pumps or compressors, because the speed of

shaft rotation is very low, there is no need to have a mechanical seal, and possibly stuffing box and packing is enough. The other example is seal-less pumps. In seal-less pumps, there is no shaft that penetrates the pump casing. The rotation of the shaft is done by magnetic induction. Mechanical seals could also be needed for mixers because there is again a shaft that’s penetrating an enclosed area. The mixers may need also to have a seal flush plan, which may not necessarily be the same seal flush plan of pumps.

The above explanation is about “liquid lubricated mechanical seal.” There is another type of mechanical seal named “gas lubricated mechanical seal” which will be discussed in gas mover section.

10.6.11 Merging Pumps

Merged pumps are basically multi-service pumps that may have multiple destination or multiple sources. They are attractive options from an economical viewpoint but they are not always easy to implement.

Both multiple destination and multiple source pumps need care in the design and P&ID development.

Multiple source pumps are less common than multiple destination pumps.

When a pump gets suction from more than one source, it is hard to predict each time which source is feeding the pump, and this is the problem. The pressure of the source may change during the operation and each stream from the source to the suction of pump could be “weaker” or “stronger” than the other stream and push the other stream away or vice versa. One solution is to use a block valve (manual or remotely operated) to change the source to whatever the intention is. Otherwise, a precise design is needed. In summary, simultaneous multiple sources are difficult to implement while non-simultaneous multiple sources are easy to implement.

This again is the case for pumps with multiple destinations. If streams are non-simultaneous, they can be easily managed by placing block valves (manual or remotely operated) to direct the flow to whatever destination. However, if the destinations need the flow simultaneously, the design is not easy. One aspect of the design is to make sure piping to “all” destinations have the same design pressure equal to the dead head pressure of the pump.

There is another non-rare case that a centrifugal pump is merged with a mixer! (Figure 10.28).

In this arrangement a pipe is branched off from the discharge of the centrifugal pump and is routed back to a container in need of mixing. Additional systems may be needed on this pipe. For example a designer may place an RO on this pipe to decrease the pressure.

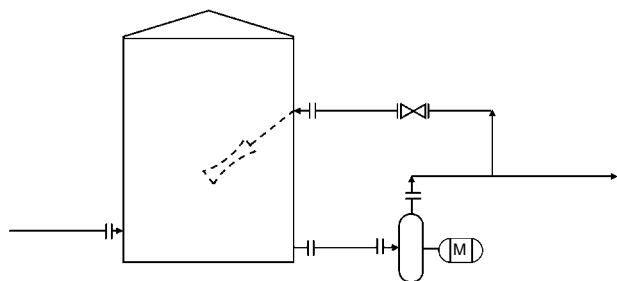


Figure 10.28 Centrifugal pump also working as a mixer.

However it should be realized that in this arrangement, the fluctuation in the pump service impacts the mixing and its effectiveness.

10.7 Liquid Movers: PD Pumps

Positive displacement pumps are not as common as centrifugal pumps in industry. When there is a need for a pump in industry the first choice is a centrifugal pump by default. PD pumps are not very user friendly and operators generally don't like them. In similar capacities they are more expensive. However, they have several advantages over centrifugal pumps. They have the least sensitivity toward cavitation. This is the most important advantage of PD pumps over centrifugal pumps. The other advantage is that they are not sensitive toward a low flow rate. They are also good if the viscosity of fluid is high or very high and/or when a very high discharge pressure is needed. There are two main classes of PD pumps: reciprocating pumps and rotary pumps.

Even though PD pumps provide a lower flow rate than centrifugal pumps in comparison with reciprocating pumps and rotary pumps, reciprocating pumps are generally used when a higher flow rate than a rotary pump is needed. However, the reciprocating pumps generate lower discharge pressure than rotary pumps.

One weakness of reciprocating pumps is that they create a pulsating flow. This pulsation is because of the forward and backward movement of the pump's motive member (Figure 10.29). For example, when you are

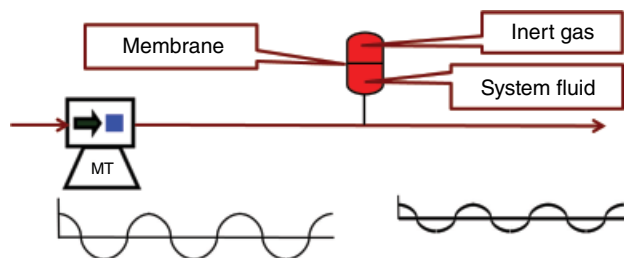


Figure 10.29 Pulsation in reciprocating pumps.

buying a reciprocating pump whose operating pressure is 500 psi, the pressure that is sensed by the elements installed on the discharge side of the pump could be between 400 and 600 psi. This is a very important point when you decide to install sensors on the discharge side of reciprocating pumps. The pulsation can be mitigated by installing a “pulsation dampener” on the discharge side of the reciprocating pump; however, the issue still exists in the pipe upstream of the pulsation dampener.

The issue of pulsation also exists on the suction side of reciprocating pumps, and a pulsation dampener is sometimes needed on the suction side of them too.

Rotary pumps can work in very high viscosity services and/or when a high discharge pressure is required. Actually, amongst the different types of pumps, rotary pumps can generate the highest discharge pressure but they have the lowest capacity. One famous application for rotary pumps is oil pumps in cars. In cars, the circulation of oil is necessary for lubrication and cooling purposes, and usually a pump is used for this. Because the flow rate is very small and at the same time the pressure needs to be high to be able to send the oil through the capillary pipes around the cylinder blocks, the best choice of pump here is a rotary type pump.

Rotary pumps are available in different types including: screw pump, lobe pump, and peristaltic pump (hose pump).

The various types of PD pumps are shown in Figure 10.30.

However, it is important to know that not all PD pumps can be clearly classified in one of these two classes of reciprocating pumps or rotary pumps. One exception is progressive cavity pumps. Progressive cavity pumps are basically a hybrid type of PD pump. These pumps are basically a mixture of reciprocating pumps and rotary pumps. As these pumps were primarily invented by a French engineer named Mono these pumps were named by people as the Mono pump.

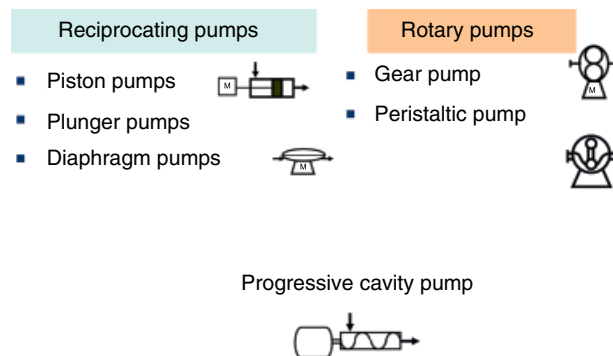


Figure 10.30 Symbols of various PD pumps.

10.7.1 PD Pump P&ID Piping

As the P&ID arrangements for a reciprocating pump are very different from rotary pumps the P&ID arrangement for these two classes of PD pumps are discussed separately.

10.7.1.1 Reciprocating Pumps P&ID Piping

The most important feature of reciprocating pumps is the generation of pulsation. Reciprocating pumps generate pulsation mainly in their discharge site and also in their suction site. If this pulsation is not desirable it should be mitigated or dampened. Generally speaking pulsation is more problematic in the discharge site of reciprocating pumps. In such cases, a device named a “pulsation dampener” should be installed in the discharge site of a reciprocating pump. It is not very common to see a pulsation dampener in the suction site of a reciprocating pump, but it has been known. We need to put a pulsation dampener in the discharge site of a reciprocating pump if we cannot afford pulsation. For example, if a reciprocating pump is pumping liquid into a huge tank, pulsation is possibly not a big deal and there is possibly no need to install a pulsation dampener in the discharge site of the reciprocating pump. However, it is very important to know that without installing a pulsation dampener all the elements in the discharge site of a reciprocating pump should be rated based on the maximum pressure that can be generated on the top of the pressure pulse by the reciprocating pump. This means the pressure rating of the pipe should be based on the highest pressure generated by the reciprocating pump at its summit of pulse. This is also the case for the other items like pressure and temperature gauges, etc. If a PSV needs be installed in the discharge site of a reciprocating pump without a pulse dampener special care should be taken to decide on the PSV set pressure otherwise the PSV will pop up frequently.

It is important to know that the maximum discharge pressure on the top of pulse summit is not always reported by the pump manufacturer and sometimes this number needs to be requested if the intention is not to place a pulsation dampener in the discharge site of a reciprocating pump.

A pulsation dampener is basically a small capsule that is connected to the pipe. This small capsule is internally split into two spaces by an elastic membrane. The top portion of the elastic membrane is filled with an inert pressurized gas similar to nitrogen. The bottom portion of the capsule is filled with the discharging liquid. This structure of a pulsation dampener helps to dampen the pulse by pressurized gas.

There are some other items that should be considered when developing the P&ID for a reciprocating pump.

A reducer or enlarger may be needed to connect flanges of a reciprocating pump to the piping. The reciprocating pumps flanges could be smaller or larger than pipe flanges in suction and discharge sites. This means there could be a need for a reducer or enlarger or nothing in the suction site of reciprocating pumps and it is also the case for the discharge site. This is one important difference between centrifugal pumps and reciprocating pumps. If you remember, in the centrifugal pump case, in the suction site only a reducer is acceptable and in the discharge site only an enlarger is acceptable.

A strainer could be installed in the suction site of a reciprocating pump. Installing a suction strainer for a reciprocating pump could be more important than for a centrifugal pump. The reason is that reciprocating pumps have more tight clearances that can be plugged by a suspended solid in the pumping liquid.

A pressure safety device should always be installed in the discharge site of reciprocating pumps. This is another important difference between centrifugal pump and reciprocating pump P&ID drawing. Whether the pressure safety device is a PSV or a rupture disk (RD) depends on the situation. In Chapter 12 it will be discussed which pressure safety device is suitable in each specific situation. The requirement of a pressure safety device is because of the type of reciprocating pump curve. As the reciprocating pump curve is a fairly straight and vertical line in the case of inadvertently closing a valve in the discharge site of a reciprocating pump (for example discharge isolation valve) the pressure starts to increase and never stops until the pump casing and/or the discharge side piping is ruptured. The intention of installing a pressure safety device not only protects the reciprocating pump casing but also protects the piping on the discharge side of the reciprocating pump. A pressure safety device in the discharge of a reciprocating pump never ever can be eliminated.

The lack of a discharge check valve is another difference between the reciprocating pump arrangement and centrifugal pump arrangement in a P&ID context. In centrifugal pump a check valve is always necessary to be placed in the discharge site of the pump; however, such a thing doesn't need to be placed for reciprocating pumps. The reason that reciprocating pumps don't need a check valve in their discharge side is because they have an internal check valve. The internal check valve in reciprocating pumps prevents backflow to the reciprocating pump in the case of sudden pump trip.

Another difference between reciprocating pumps and centrifugal pumps is the lack of a minimum flow protection pipe in reciprocating pump arrangements. You remember that it was mentioned in Section 10.6.1 that in the majority of cases a minimum flow protection pipe is needed to put around a centrifugal pump to protect it

from very low flow rates during the reduced flow rate conditions in a processed plant. However, such a thing doesn't need to be placed in reciprocating pump arrangements. The reason is not that there won't be any minimum flow experienced in a process plant. The reason is that reciprocating pumps are generally insensitive toward the low flow rate. This means reciprocating pumps still work fine even when the flow rate to them is very low. However, it should be mentioned that it is not the case that reciprocating pumps work fine even when the flow rate is very tiny in comparison to the size of the reciprocating pumps. When the flow rate is very tiny they start to lose their efficiency. However, this doesn't force us to provide a minimum flow protection pipe for them as this very low flow rate is much lower than the low flow rate that can be experienced in process plants.

A typical P&ID representation of reciprocating pumps is shown in Figure 10.31.

10.7.1.2 Rotary Pumps P&ID Piping

For rotary pumps, there is no need to install a pulsation dampener in the discharge site or suction site of rotary pumps. This is because rotary pumps don't generate pulsation even though they are part of positive displacement pumps.

Similar to reciprocating pumps, in rotary pumps the flange sizes in the suction site and discharge site could be the same, smaller, or larger than pipe flanges. Therefore, having a reducer, and enlarger, or nothing in the suction and/or discharge sites of rotary pumps would be acceptable.

Rotary pumps need a pressure safety device on their discharge side. This is because of the same logic as for

reciprocating pumps. However, some companies decide to take a less conservation approach and do not install a pressure safety device on the discharge side of rotary pumps. Their logic is that because rotary pumps have an internal pressure safety device they don't really need another external pressure safety device on their discharge site. However, the companies that are more conservative believe that the internal pressure safety device of rotary pumps cannot relieve us from putting another pressure safety device on the discharge side of rotary pumps. The reason is that the set point of the internal pressure safety device cannot be fully in the control of the designer and also that the pressure safety device is only to protect the rotary pump and not necessarily the discharge piping. Therefore, it makes sense to disregard the internal pressure safety device and put another pressure safety device on the discharge site of rotary pumps.

There is a need to put a check valve in the discharge site of rotary pumps. This is similar to centrifugal pumps and dissimilar to reciprocating pumps.

There is no need for a minimum flow protection pipe for rotary pumps. The logic is the same as for reciprocating pumps.

There are some cases where a spillback pipe between the discharge side of a PD pump can be supplied with an isolation valve. This pipe loop is very similar to a minimum flow protection pipe; however, the pipe size is different and also this has an isolation valve like a gate valve rather than what we have in a minimum flow protection pipe, which may have a control valve. The purpose of this pipe loop is to facilitate a start-up of PD pumps.

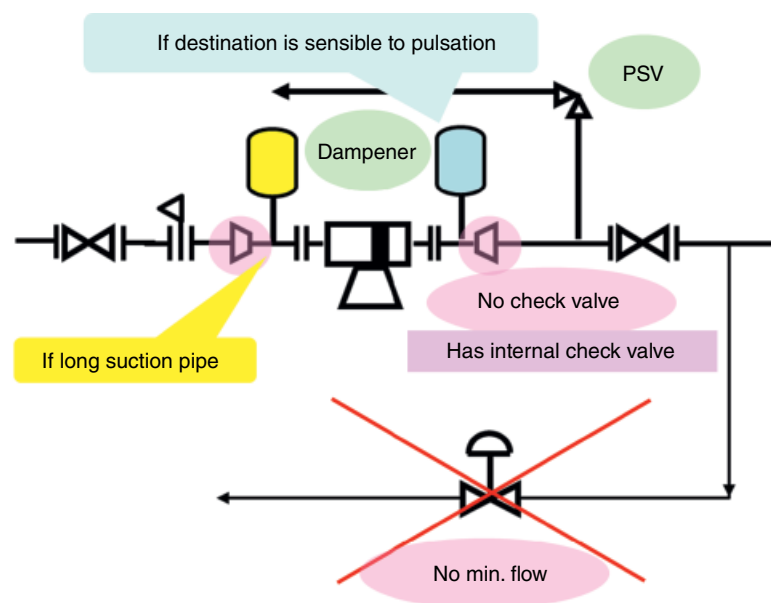


Figure 10.31 P&ID arrangement of a reciprocating pump.

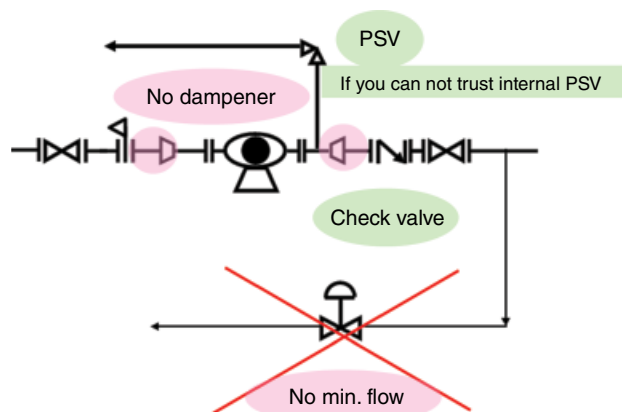


Figure 10.32 P&ID arrangement of a rotary pump.

A typical P&ID representation of rotary pumps is shown in Figure 10.32.

10.7.2 PD Pump Arrangements

PD pumps could be placed in parallel to provide a spare pump. However, having two (or more) PD pumps functioning at the same time is not very common. Less common is using PD pumps in series.

When two (or more) PD pumps are placed in a parallel arrangement they may need additional considerations during P&ID development. Some companies put another PSV for each PD pump in parallel to protect the suction side of them from being over-pressurized by the operating PD pump.

If the discharge isolation valve of the spare PD pump is negligently left open, the operating PD pump pressurizes the spare pump, both on the discharge side and suction side. Another PSV around the suction isolation valve protects the suction side of the PD pump (Figure 10.33).

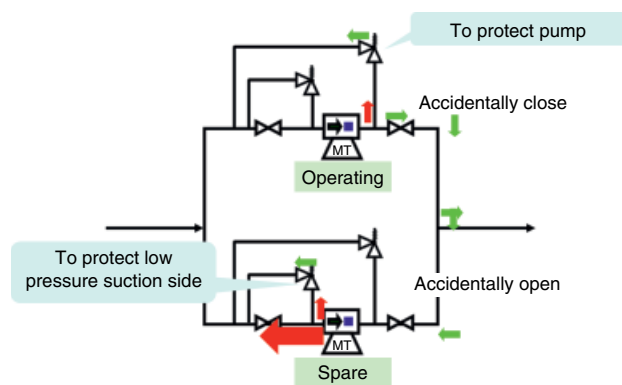


Figure 10.33 Additional PSVs for PD pumps in parallel.

10.7.3 Merging PD Pumps

Similar to centrifugal pumps, PD pumps may also be used in multi-service applications. However such applications are much less common than centrifugal pumps.

It is generally recommended to use one PD pump for one suction source and one discharge destination.

10.7.4 Tying Together Dissimilar Pumps

Here we are talking about specific requirements when dissimilar pumps are tied together. Such practice is not a very good practice but sometimes can be done after considering all the requirements.

For a parallel arrangement of dissimilar pumps (Figure 10.34) there could be different reasons for paralleling, similar to centrifugal pumps. However, it is not very common to see such arrangements in process plants. One application could be when the viscosity of the service liquid is changing a lot and two different pumps should be ready to be able to handle liquids with different viscosities through two different pumps. The other cases could be when the main pump is the centrifugal pump and a small PD pump is used just to keep the system pressurized when the centrifugal pump is off.

For a series arrangement of dissimilar pumps one rule of thumb can be memorized: “using a PF pump as the booster pump of another pump is not good practice.”

Different cases of this rule of thumb and its applicability can be seen in Figure 10.35.

10.7.5 PD Pump Drives

The most common type of drives for PD pumps is electric motors. This is similar to centrifugal pumps.

For reciprocating type PD pumps there are, however, some other options available.

One option is air-operated drives. In air-operated drives a source of air, like plant air, is used to generate a reciprocating movement. This reciprocating shaft is connected to the pump shaft to operate it. Such pumps may be named “air-driven pumps.” This type of drive is very common for diaphragm pumps. It should be

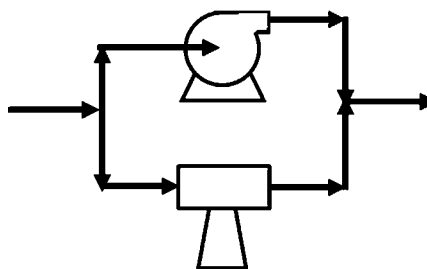


Figure 10.34 Dissimilar pumps in parallel.

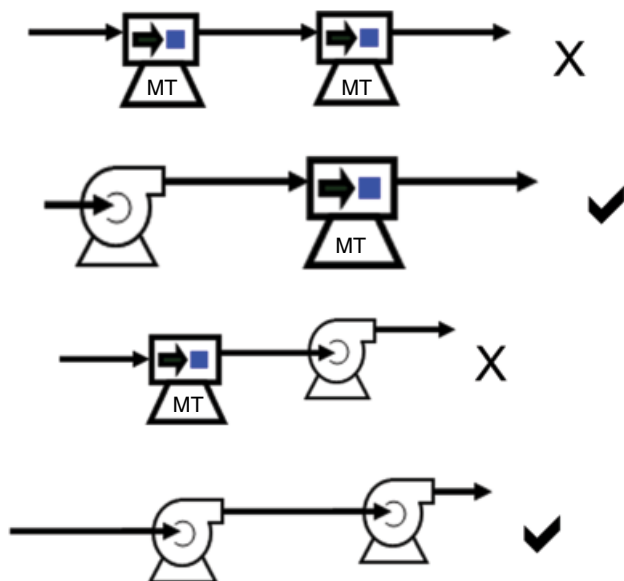


Figure 10.35 Different arrangements of dissimilar pumps in series.

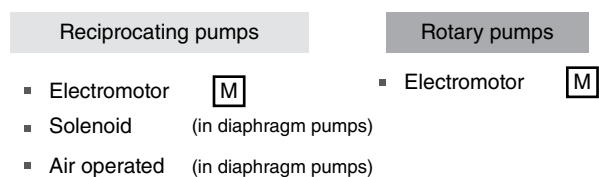


Figure 10.36 PD pump drives.

mentioned that the maximum discharge pressure of these pumps is limited by the pressure of the air source.

The other option is used in “solenoid driven pumps.” In “solenoid driven pumps” the drive is a stroking shaft driven by a repeatedly magnetized solenoid.

Companies may or may not decide to show the pump drives. Figure 10.36 shows the P&ID representation of different drives.

Even if it is decided to not show pump drives, it is very difficult to not show air-operated drives as they are a type of process elements.

10.7.6 Sealing Systems for PD Pumps

PD pumps similar to centrifugal pumps may need a sealing system. However, their system is not standardized and is designed by the vendor. If the rpm of the shaft is low enough, the designer may decide to use packing rather than a mechanical seal and then the need for a sealing system is eliminated.

In a reciprocating pump the shaft is reciprocating rather than rotating and the sealing concept could be totally different.

10.7.7 Metering Pumps (Dosing Pumps)

Metering pumps or dosing pumps are pumps that are able to generate flow with adequate accuracy within a reasonable range of upstream and downstream pressures.

Because obviously changing upstream and downstream pressures will change the flow rate of centrifugal pumps they cannot be categorized as metering pumps. Metering pumps are generally referred to as PD pumps. The majority of metering pumps are reciprocating type PD pumps because they have more flexibility in controlling them.

Dosing pumps may be controlled through VSD and also change in stroke length (refer to Chapter 15).

As the amount of injected chemical is generally important, the injecting flow rate should be checked occasionally by operators. In the majority of cases this is done by providing a drawdown (calibration) column on the suction side of the pump. The rounding operator closes the blocking valve upstream of the calibration column and allows the pump to get suction from the liquid inside of the column (instead of the upstream container) and measures the time for a specific drop in the liquid of the calibration column to calculate real flow rate (Figure 10.37).

In some cases it is very important to install a back-pressure regulator. This is to mitigate one condition that may cause uncontrolled injection of chemical to the host stream. If the pressure of the host stream (which could be fluctuating) goes below the suction pressure of the

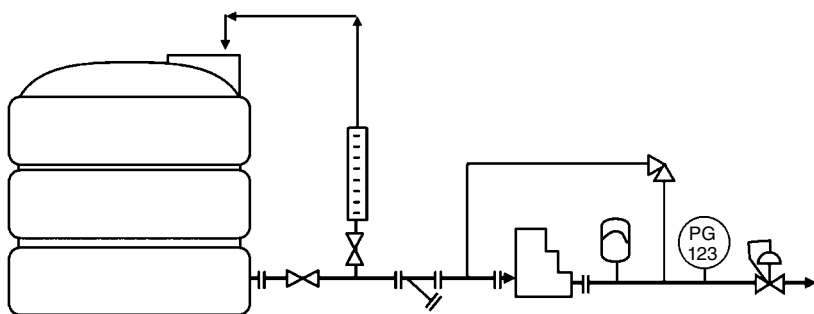


Figure 10.37 P&ID representation of a dosing pumps.

dosing pump, the host stream “sucks” the chemical with an unknown flow rate. To make sure that the discharge pressure is always higher than the suction pressure, a back-pressure regulator needs to be installed.

10.7.8 Liquid Transfer – Summary

Where a liquid needs to be transferred and the pressure to be compensated for (or “head” in meter units) difference is less than 50 kPag, the liquid flow is generated only by elevating the source point (and/or lowering the destination point) in a way to initiate a gravity flow. Generally it doesn’t make sense to use any type of mechanical system like a pump to transfer the liquid where such small head is needed.

For higher required head a pump is the default choice. In the majority of cases a pump, either centrifugal or positive displacement, is used for liquid transfer.

The other cases are when a liquid transfer is needed but not continuously and intermittently and very sparingly. In such cases it is possible that using a pump cannot easily be justified. In such cases liquid transfer with the help of pressurized gas could be a good option.

One system that can work for this intention is the “blow case” (Figure 10.38).

The main piece of equipment in a blow case system is a container or vessel. The liquid that needs to be transferred is collected in this vessel. After a while, when enough liquid is collected in the vessel, the liquid is pushed out of the vessel by gas pressure. Thus, this vessel has an inlet pipe for liquid and an outlet pipe for liquid and each of them has a valve on it. The vessel could be equipped with a level gauge or a level indicator.

There is another incoming stream to this vessel and it is the motivating gas stream. This pipe also has a valve on it.

During the collection of liquid in the vessel all valves are closed except the valve on the incoming liquid. When it is decided to transfer the collected liquid, the incoming liquid valves will be closed then the incoming gas will be

opened to pressurize the system, and then the outlet liquid gas is opened to transfer the liquid.

In some designs the operation is automated by replacing the level indicator with a level controller and all the manual valves with switching valves and through a process interlock system.

If the operation is manual the operator needs to make sure that the pressurized gas doesn’t sweep away from the vessel and go to the liquid destination. This can be done by closing the liquid outlet valve before completely emptying the vessel. However, as “gas bypassing” may happen, the pressurizing gas should be selected to create fewer problems in the destination container. For example if the destination container is a natural blanketed tank, the pressurizing gas could be natural gas, and if it is a non-blanketed tank, the pressurizing gas could be plant air.

A blow case’s structure is similar to “pneumatic ejector pumps.”

The other example of using gas to move a liquid is in off-loading of non-innocent liquids from a tanker (for example) to a tank. The transfer of hydrochloric acid from a tanker to a tank needs an acid resistant centrifugal pump, which could be expensive. A source of gas can be used to be applied to the surface of the liquid acid in the tanker to push it from the tanker to the storage tank. The gas could be plant air or it could be compressed air generated by a compressor installed on the truck.

Table 10.9 outlines some rules of thumb regarding the selection of liquid movers.

10.7.9 Pumps: Duty Other than Pumping!

A pump can be considered as pump + valve in some applications.

A centrifugal pump with VSD could be used as a pump plus a control valve on a flow loop.

A PD pump with VSD could be used as pump plus a control valve on a pressure loop with a pressure sensor upstream of the valve (or a back-pressure regulator).

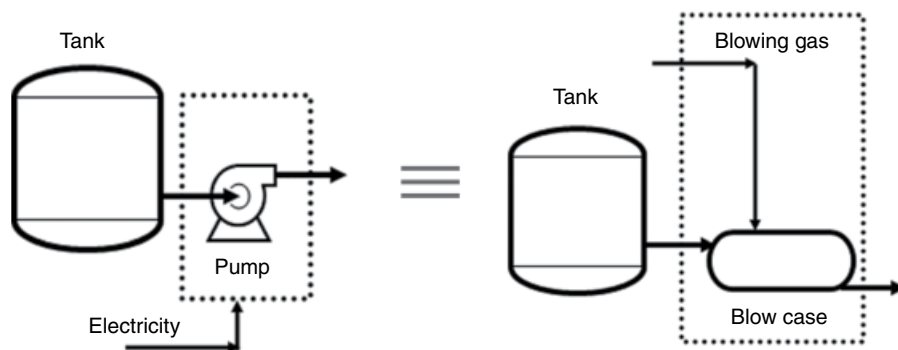


Figure 10.38 Transferring liquids by gas pressure through a blow case.

Table 10.9 Rules of thumb for the selection of liquid movers.

Liquid mover type	Driving force	When?
Gravity flow	Difference in elevation	Where less than 50 kPag is needed to move liquid
Gas pushing transfer	Gas pressure	<ul style="list-style-type: none">• Where less frequent intermittent operation is needed• Where the transferring liquid is aggressive
Dynamic pump	Axial pump	When flow rate is huge
	Centrifugal pump	Default choice
PD pump	Reciprocating	For low flow rates (say less than 5 m ³ h ⁻¹), and high viscosity liquids (say more than 200 cP)
	Rotary	For cases more stringent than the cases for reciprocating pumps

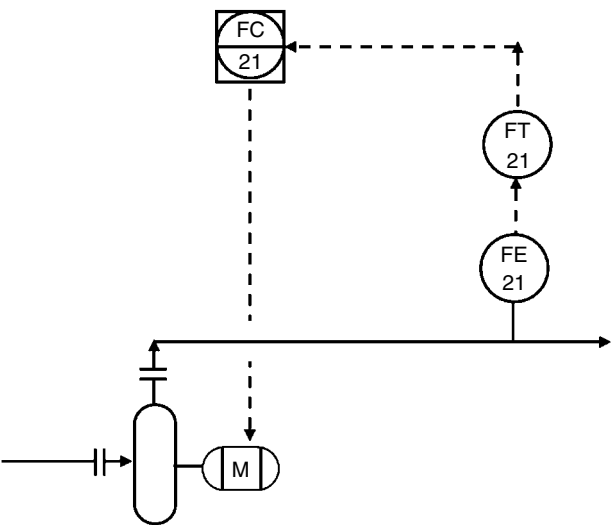


Figure 10.39 Pump is role of pump plus valve.

It is not common to eliminate the control valve because of the existence of the pump, but there are cases that such thing can be done. For example when the pumping liquid is very viscose or the pumping liquid is very abrasive.

This concept is shown in Figure 10.39.

10.8 Gas Movers: Fans, Blowers, Compressors

For gases, there are three main types of movers: fans, blowers, and compressors. The first two movers (fans and blowers) are basically only for transfer or mobilization of the gases but the third type, the compressor, is used for mobilization and pressurization of the gas. If the only purpose is transferring a gas from point A to point B and there is no need to generate high pressure in the pipe, then a fan or a blower is enough. However, if, in addition to transferring gas from point A to point B, the

gas needs to be pressurized for the purposes of a downstream unit of operation or process, then a compressor needs to be used. However it should be mentioned that there are cases where the intention is only transferring gas but the pressurization is also implemented to reduce the size of the pipes.

Axial gas movers are not very popular and they have very limited applications. Their main element is a propeller, similar to a ceiling fan. One of their common applications is in “air coolers,” to push the air across the tube bundle to cool down the fluid flowing inside the tubes. Axial flow gas movers are used for huge flow rates (say more than 300 A m³ s⁻¹) or higher.

Centrifugal gas movers are used mainly in large industries dealing with large flow rates of gases, up to say 200 A m³ s⁻¹. The famous examples are gas processing industries, including natural gas treatment and natural gas liquids (NGL) industries.

Other than those industries, the applications for centrifugal gas movers are limited. This is because centrifugal fans/blowers or compressors have huge capacities (flow rates), which are generally larger than what is needed in most industries.

The majority of gas movers in industry are the PD type. They are generally for capacities less than 5 A m³ s⁻¹.

Regarding the weaknesses of centrifugal compressors, we may think that centrifugal compressors have the same weaknesses as centrifugal pumps, and this assumption is not far off. As we saw, centrifugal pumps have two big problems: cavitation and low flow intolerance; however, centrifugal compressors have only one problem: “low flow intolerance.” The reason is obvious; cavitation is only an issue when we are dealing with liquids.

10.8.1 Low Flow Intolerance and Anti-Surge Systems

As was mentioned, centrifugal compressors, similar to centrifugal pumps, are intolerant toward low flow operations. A system must be implemented around a centrifugal

compressor to address this problem. The name of that system is the “anti-surge system.”

Table 10.10 summarizes the similarities and differences between minimum flow in centrifugal pumps and compressors.

Reciprocating gas movers: reciprocating compressors are the type of gas mover corresponding to PD pumps. They are used for higher required discharge pressures (generally). Their capacity (flow rate) is much smaller than that of centrifugal gas movers and if the required flow rate is high, then multiple units in parallel should be used.

Rotary gas movers: rotary gas movers are used for small flow rates. If a gas mover is required to be installed on a truck or small trailer in a portable system, the choice is generally a rotary gas mover.

They come in different types and names, including lobe compressors and screw compressors.

From the other side, gas movers can be classified as fans, blowers, and compressors. Then a matrix, similar to one shown in Table 10.11, can be developed to introduce a full range of gas movers.

10.8.2 P&ID Development of Gas Movers

One may decide to develop the P&ID of gas movers based on the corresponding liquid movers. This is not a

bad decision; however, there are several differences between gas movers and their corresponding liquid movers. They are:

- As gas movers are working on gases that have higher volume than liquids per unit of mass, they are generally bigger than their corresponding liquid movers.
- The bigger sizes of gas movers may force the manufacturers and P&ID developer to consider some points that were not an issue in liquid movers.
- The work required to be done to transfer and/or pressurize gases is higher in gas movers in comparison to work needed for transferring liquid.
- Compression of gases generally makes the gas hot and hotter. Thus the gas needs to be cooled down even in the middle of pressurizing. This can be done by using a multi-stage compressor with intercooling.
- A cooling system and a lubrication system are two systems that are seen in the majority of compressors (but not necessarily in fans or blowers).
- Gas movers are sensitive to the existence of liquid droplets in to-be-pressurized gas. Then liquid bubbles need to be removed upstream of the gas mover through scrubbers. It may be said that there is a similar issue in liquid movers; liquid movers are sensitive toward the existence of gas/vapor bubbles. However the chance of having a gas stream with non-tolerable liquid content

Table 10.10 Minimum flow in centrifugal pumps and compressors.

Parameter	Liquid mover: centrifugal pumps	Gas mover: centrifugal fans/blowers/compressors
Issue	Low flow instability	Surge issue
Result	Unstable operation (recirculation instability), overheating	Very quick and frequent change in the output flow rate of the gas mover
Goal	The flow should be always higher than the “minimum continuous stable flow” (MCSF) of the pump.	The flow should be always higher than the “minimum flow” of the compressor.
Susceptible fluid movers	More problematic for larger pumps and pumps with lower $NPSH_R$	Mainly an issue for compressors and sometimes blowers. Not an issue for fans.
Protection	Minimum flow recirculation pipe (or spillback pipe) system	Anti-surge system

Table 10.11 Summary of various gas movers.

	Dynamic type		Positive displacement type	
	Centrifugal	Axial	Reciprocating	Rotary
Fan Pd < 20 kPag	Up to 20 kPag	Up to 5 kPag	Not available	Not available
Blower Pd < 200 kPag	Available	Available	Up to 200 kPag	Up to 100 kPag
Compressor Pd > 200 kPag	Up to 100,000 kPag	Not common	Up to 1,000,000 kPag	Up to 1,000 kPag
Comment	Medium to big size	For innocent gases	Small to medium size	Low capacity

(wet gas) is higher than for liquids with non-tolerable gas content.

- As the gas movers are large, their moving elements may have huge momentum and in the cases of a quick trip in the system, they may move for a while even after shutdown. Therefore there should be some systems to take care of gas movers in those situations. Some of the required actions could be implemented as part of the interlock system. Interlock systems will be discussed in Chapter 16.
- If the to-be-pressurized gas is flammable, more caution should be taken in design. This is mainly because by pressurizing a gas, it will be hot.
- Compressors may be placed in an enclosed or semi-enclosed building for different reasons. One reason could be the sound level around the compressor. The other reason is to provide a dedicated space for compressors for ease of inspection by operators. The “compressor building” could be a multi-story structure for the ease of inspection and maintenance.

10.8.3 Gas Mover Drives

The same drives of centrifugal pumps are available for gas movers too. Electric motors, turbines (steam driven and gas turbines), and internal combustion motors are some of them.

10.8.4 Auxiliary Systems Around Fluid Movers

Auxiliary systems around fluid movers could be supporting systems for different components of the fluid mover. A fluid mover consists of drive section and fluid mover section. A “connection” attaches these components together (Figure 10.40).

The “connection” could be shaft, gear box, crankshaft or a combination.

Each of the above components may need some type of auxiliary system. Some of these auxiliary systems are listed in Table 10.12.

The seal type for the gas movers could be the same type of seal types for liquid movers. The liquid mover seal type was discussed before and is a “wet type sealing.” For a long time it was assumed that the only type of seal for compressors are the wet type. But later it was recognized that even a gas could be used as a sealing fluid. Then a newer type of sealing system, the dry gas seal, came on the scene.

In a dry gas sealing system an inert gas like nitrogen gas can be used as the sealing gas.



Figure 10.40 Fluid mover and drive as a pair.

Table 10.12 Fluid mover auxiliary systems.

Components	Type	Required auxiliary system
Drive	Electric motor	<ul style="list-style-type: none"> • Bearings may need cooling • The majority of motor blocks are generally air cooled and no specific system is required. However some of them need forced air cooling and air should be cooled (by air or water) and recirculated. Some huge electric motors could be oil cooled, then a cooling system is needed
	Internal combustion engine (diesel, gas)	<ul style="list-style-type: none"> • Pistons need lubrication • The motor block needs cooling
	Air operated drive	<ul style="list-style-type: none"> • No specific auxiliary system is required. Only plant air should be piped
	Solenoid operated drive	<ul style="list-style-type: none"> • No specific auxiliary system is required. Only DC electricity is required
	Steam turbine	<ul style="list-style-type: none"> • The block may need cooling
	Gas turbine	<ul style="list-style-type: none"> • The block needs cooling
Connection	Shaft	<ul style="list-style-type: none"> • The penetration point should be sealed, lubricated, cooled, and flushed
	Gear	<ul style="list-style-type: none"> • Gear box may need lubrication
	Crankshaft	<ul style="list-style-type: none"> • Crankshaft may need lubrication
Fluid mover	Liquid mover	<ul style="list-style-type: none"> • Bearings may need cooling
	Gas mover	<ul style="list-style-type: none"> • (As mentioned) penetration points should be sealed, cooled, and flushed • In compressors the block should be kept cooled but this is done by the intercooler shown on the main process P&ID and not the auxiliary P&ID

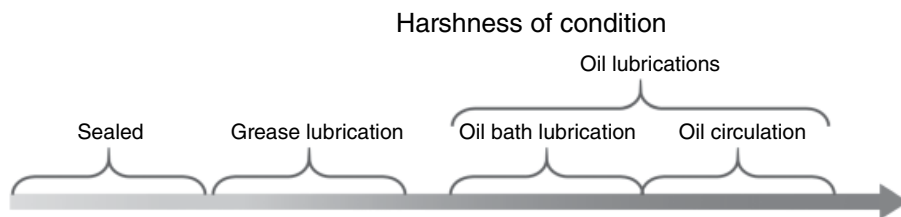


Figure 10.41 Various lubrication methods depending on the harshness of conditions.

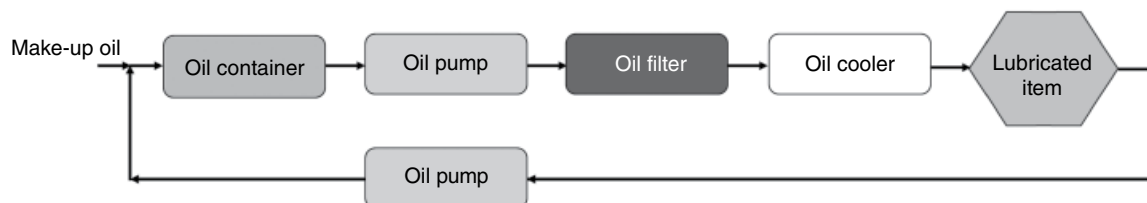


Figure 10.42 Block flow diagram of a recirculating oil system.

There are different types of lubrication systems available. Their application depends of the harshness of conditions and also economical constraints. Figure 10.41 shows a few of them.

In oil bath types, there is an oil bath and the element of interest for lubrication submerged in it or oil get transferred to (or splashed on) it through simple parts like “oil rings” or spoon-type elements.

Recirculating oil type lubrication is a more complicated system. A block flow diagram of a recirculating oil system is shown in Figure 10.42.

The oil container (which may be named an oil reservoir or oil sump) can be equipped with a heater if there is a chance of very low ambient temperature that may “set” the oil.

The return oil pump may not be needed in all designs. In some designs the “return oil” can flow to the oil container by gravity rather than through a pump. The pumps in lubricating oils are mainly rotary pumps of types of screw or lobe pumps.

If the heat load on lubricating is not large, the oil cooler could be eliminated. The oil cooler could be an air cooler or water cooler.

Where there are several to-be-lubricated items, the provided oil after the cooler may go through the oil distribution manifold and then supplied to each to-be-lubricated item.

There are mainly two types of recirculating oil system: oil sump lubrication and oil mist lubrication systems.

In an oil sump lubricating system oil is transferred and delivered as pure lubricant and delivered in the form of oil droplets or jets.

In an oil mist lubricating system oil is transferred as pure lubricant but delivered in the form of oil mist with help and after mixing with other fluids.

There are two types of oil sump lubricating systems: wet sump and dry sump types.

For the wet sump type an attempt is made to share the elements of the lubrication system as much as possible with the to-be-lubricated system. For example, in all wet sump systems the oil container is not a dedicated and external oil tank; it is rather a portion of the to-be-lubricated system, which is voluminous and has capacity to hold the lubricating oil. In some designs the oil pump is totally independent but in others the oil pump drives from the to-be-lubricated system like a motor.

For the dry sump type the system is totally separated and external to the to-be-lubricated system. An oil tank sitting beside a to-be-lubricated system clearly shows the lubrication system is dry sump type.

10.8.5 Gas Transfer – Summary

Generally speaking mechanical gas transfer is difficult and costly. We prefer that a gas is transferred by its pressure rather than using a mechanical gas mover.

The other preference is to convert the gas to liquid and then transfer it, rather than transferring a gas! (Figure 10.43.)

There are, however, cases where transferring gases or vapors is inevitable. In those cases we use fans, blowers, or compressors.

Table 10.13 outlines some rules of thumb regarding the selection of liquid movers.

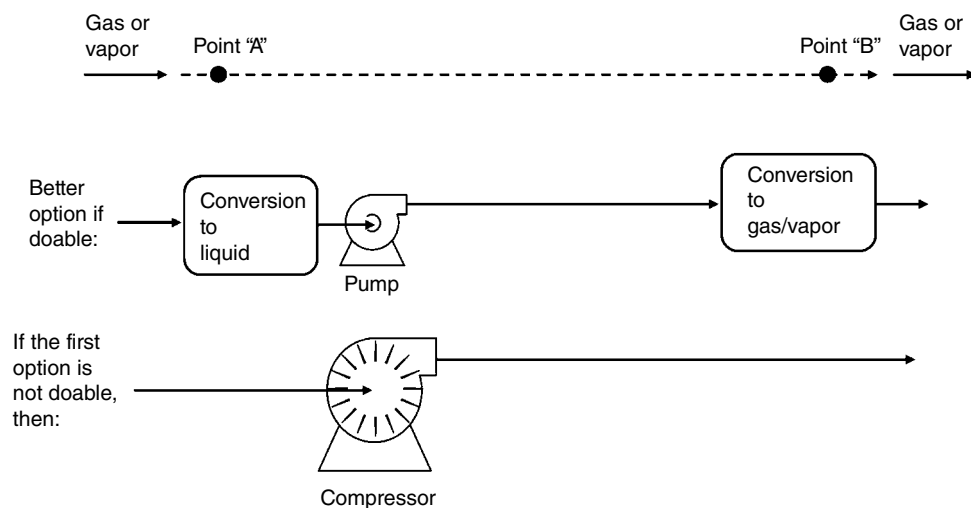


Figure 10.43 Unfavorability of gas movers.

Table 10.13 Rules of thumb for the selection of gas movers.

Gas mover type	Driving force	When?
Gas pressure transfer	Gas pressure	When enough pressure is available
Dynamic gas mover	Axial	Mainly in fans
	Centrifugal	For huge flow rates
PD gas mover	Reciprocating	Default choice for blowers and compressors
	Rotary	Common for blowers and compressors

10.9 Wrapping-up: Addressing Requirements of Fluid Movers During the Life Span

In this section we check our design to make sure we cover all the needs of fluid movers during each phase of plant life. As was discussed before, these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item.

- 1) Normal operation.
- 2) Non-routine operation (reduced capacity operation, start-up operation, upset operation, planned shutdown,

emergency shut down); these phases of plant need two components to be handled: the process component and the control/instrumentation component. The process component is discussed here but the control/instrumentation component will be discussed in Chapters 13–15.

The recirculation pipes may work as a “silver bullet” and for such cases we have already covered them. In other words, by recirculating fluid around a fluid mover, we make the conditions for the fluid mover similar to normal operation even during non-routine operation.

- 3) Inspection and maintenance: general consideration regarding inspection and maintenance of all items is covered in Chapter 8. Here, however, we cover the specific requirements.

The small liquid movers may need a few items for inspection and maintenance. However, large liquid movers and compressors may have a bunch of points for inspection. The minimum inspection points for pumps could be oil level and electrical current (amperage) reading, none of which have a footprint on P&IDs. Larger fluid movers may need temperature sensors and vibration sensors in various locations.

- 4) Operability in the absence of one item: generally a spare is provided for fluid movers. The exceptions could be huge and expensive compressors that we may not be able to afford to provide a spare for.

Reference

- 1 Woodside O. Protect centrifugal pumps from low flow, chemical engineering progress, June 1995, p.53.

11

Heat Transfer Units

11.1 Introduction

There are mainly two types of heat transfer units: heat exchangers and furnaces (fired heaters).

Often, stream temperatures need to be increased or decreased in a plant in order to satisfy process requirements.

You may say: “This is easy; I have done this a lot by putting a kettle of water on the stove.”

You are right, but in industry there are better ways to do this. In industry, changing the temperature of a fluid using an external source of energy, such as fire or an electric heater, is the most expensive method of changing the temperature, and is used as a last resort.

11.2 Main Types of Heat Transfer Units

When the temperature of a stream requires increasing or decreasing, the first choice is putting it in contact with other higher or lower temperature streams. This can be done in a heat exchanger.

In heat exchangers, the temperature of a stream is changed by putting the stream in contact with other streams to give it an opportunity for heat transfer.

When we talk about a “contacting” fluid stream, the contacting is indirect, meaning the fluids are not mixed together.

If the temperature of a stream needs to be increased, then it should be sent to a heat exchanger and placed in contact with a heating medium or a hot fluid, such as a hot process stream, steam, or hot glycol.

If the intention is to decrease the temperature of the target stream, it should be sent to a heat exchanger with a cold medium, such as cold process stream, ambient air, cooling water, or refrigerants.

If a suitable “process stream” can be found amongst available streams in the plant, we use it as the first choice. Otherwise, we may have to use “heat transfer media utility streams.”

“Utility streams” will be discussed in Chapter 17; however, they can be briefly defined as auxiliary streams needed by the main process in plant. A group of utility streams are “heat transfer media utility streams,” or the streams used to heat up or cool down the process streams. Streams like steam, hot glycol, or synthetic heat transfer fluids like Therminol[®] are hot utility streams that are used for heating up the process streams. Streams like cooling water, ambient air, and cold glycol are cold utility streams that are used for cooling down the process streams.

If a suitable stream doesn’t exist for this purpose, or the temperature of a stream needs to be increased excessively, the other option is to use a furnace/fired heater, which is equipment that changes the temperature of a stream with the help of a primary source of energy, such as fire. There are other popular names for furnaces in industry, such as heater or boiler. If a furnace is used to increase the temperature of water (generating vapor), then the furnace is called a boiler.

However, increasing the temperature of a stream using a furnace is a very expensive practice, and in industry we try to do this temperature change using heat exchangers as much as possible, and not furnaces. That is why in a plant there may be many heat exchangers, but only one or two furnaces.

This concept is shown in Figure 11.1.

When it is decided to use heat exchanger there are questions that should be answered:

- Which type of heat exchanger should be used?
- Which heat transfer fluid should be used in the heat exchanger?
- Which arrangement should be used, a single heat exchanger, or heat exchangers in series or in parallel?

All of three above questions are generally answered in the FEED stage of the project and there is nothing on the shoulders of the P&ID development engineer. However, here we briefly cover the topics as there have been some cases where such decision has been delayed even to the detail stage of project.

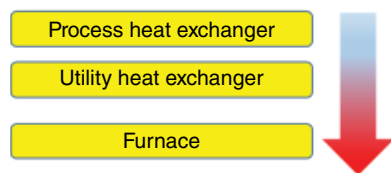


Figure 11.1 Usage of heat transfer units in process plants.

11.3 Different Types of Heat Exchangers and Their Selection

Different types of heat exchangers are used in the process industry. The reason for the different available types is the differences in indirectly contacting two fluids to each other.

For example in shell and tube heat exchangers cold and hot streams transfer the heat through the peripheral area of several tubes.

The most common type of heat exchanger is the “shell and tube” (S&T HX) type. In this type of heat exchanger, a bundle of tubes is secured on one or both sides in tube sheets and is placed inside a cylindrical body, or shell. A tube sheet is a perforated sheet that secures the tubes in its holes. One fluid is flowing through the tubes and the other fluid is flowing in the space between the outside of the tubes and the shell.

If a tube bundle comprises only one tube, this could be considered as a variation of a S&T HX and it is named a “double pipe heat exchanger.”

After shell and tube heat exchangers, possibly the most common heat exchangers are “plate and frame (P&F HX) types. P&F HXs provide channels for each stream.

In this type of heat exchanger, several plates are put together in the form of a sandwich and are secured between two “jaws.” Plates are generally in rectangular shape. The plates are separated from each other by peripheral gaskets, which provide a gap between every two adjacent plates. Hot streams and cold streams flow through these narrow gaps, or channels.

The last type of heat exchanger is the spiral type. To visualize this type of heat exchanger, consider a P&F HX with a larger-than-usual plate size. If someone rotates and wraps this “sandwich” around a core, then a spiral heat exchanger is obtained. Spiral heat exchangers

Table 11.1 Rule of thumbs for selecting heat exchanger types.

Heat exchanger type		When?
Shell and tube (S&T) heat exchanger	Fixed head	Default choice but applicable if desired temperature is less than 50–60°C (or less than 80–90°C with expansion ring)
	Floating head	Where the desired temperature is more than 50–60°C <ul style="list-style-type: none"> Where the shell side fluid is fouling
Double pipe heat exchanger		When the decision is S&T but the required heat duty is low
Plate and frame (P&F) heat exchanger		<ul style="list-style-type: none"> Wherever is not enough room for an S&T type but enough pressure is available Pressure and temperature cannot be severe otherwise the tolerance of the gaskets of the P&F HX will be exceeded Where the required heat duty is not certain and the modular structure of P&F helps in the future addition of plates (and increase in the heat transfer area)
	Spiral heat exchanger	For very fouling services
	Aerial cooler	When cooling down to approximately 65°C is adequate
Helical coil		For small heat duties, for example sample coolers and pump seal water heat exchangers

are the least common and the most expensive type of heat exchangers.

There are plenty of criteria that should be considered for decision on a specific type of heat exchanger. Table 11.1 summarizes some rules of thumb for the selection of heat exchangers.

Another rule of thumb uses the required heat transfer area for the selection of heat exchangers. This rule of thumb is presented in Figure 11.2.

At the end of this section one important piece of terminology needs to be discussed.

Each heat exchanger has two enclosures in contact with each other and with a common wall.

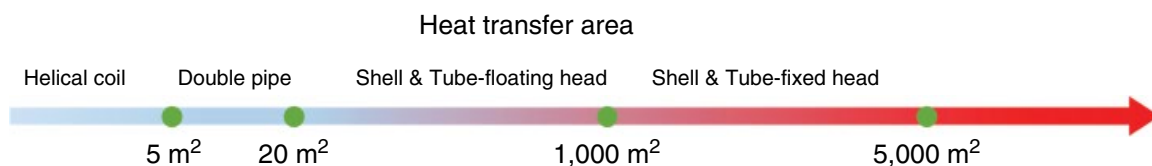


Figure 11.2 Selection of heat exchangers based on the required heat transfer area.

Table 11.2 Terminology of twin enclosures in heat exchangers.

Heat exchanger type	Enclosure Name
Shell and tube (S&T) heat exchanger	Shell side, tube side
Double pipe heat exchanger	Pipe side, annular side
Plate and frame (P&F) heat exchanger	Hot side, cold side
Spiral heat exchanger	Hot side, cold side
Aerial cooler	Tube side

Each enclosure is for each stream, one for a cooling stream and the other one for a heating stream.

To refer to each of these enclosures we have to use better terminology than “closure one of the heat exchanger” and “closure two of the heat exchanger.”

In S&T HXs, a stream flows inside of tube and the other stream flows outside of the tubes or in shell. Therefore we can name the former enclosure the tube-side and the latter the shell-side.

This applies specifically to S&T HXs. The other types of heat exchangers don’t have shells and tubes but they still have two enclosures. Table 11.2 lists the terminology for the two enclosures in different types of heat exchangers.

11.4 Different Types of Heat Transfer Fluids and Their Selection

As it was mentioned before we have two types of heat exchangers of process heat exchangers and utility heat exchangers, and process heat exchangers have higher priority for use in plants.

We can, then, say the best heat transfer fluid is the existing process fluid and then utility fluids.

Amongst cooling utility streams the best fluids are air and water, which are abundant resources. Therefore they are on the top of the list of preferred utilities. If they cannot be used the other options can be considered.

If sea water is available it could be very attractive choice for cooling.

A list of cooling streams is given in Table 11.3.

For heating purposes the fluids on the top of list are hot water and steam.

A list of heating streams is given in Table 11.4.

If the temperature of the target stream needs to be increased to more than 400 °C, the only choice is probably a fired heater.

Using steam as a source of heat in heat exchangers is very common. However there are some points regarding the usage of steam as a heat transfer medium.

Table 11.3 Utility stream choices for cooling.

Cooling streams	Application
Cooling by air in aerial cooler	When cooling down to approximately 65 °C is adequate
Cooling by “cooling water” or “cold glycol”	When cooling down to 65 °C is not enough but down to approximately 40 °C is adequate
Cooling by “chilled water”	When cooling down to 40 °C is not enough but down to approximately 20 °C is adequate
Cooling by “refrigerated water”	When cooling down to 20 °C is not enough but down to approximately 10 °C is adequate

Table 11.4 Utility stream choices for heating.

Heating streams	Application
Heating by hot water or hot glycol	When heating up to approximately 100 °C is adequate
Heating by steam or “hot glycol”	When a heating up to 100 °C is not enough but up to approximately 150 °C is adequate
Heating by non-water based hot liquids	When heating up to 150 °C is not enough but up to approximately 400 °C is adequate

The first point is that only saturated steam, and not superheated steam, can work as the heat providing fluid. If superheated steam is available, and it is intended to be used for heating purpose in a heat exchanger, it should be converted to saturated steam before usage. A superheated steam is nothing except a “gas,” but it can be converted to a heating medium through a desuperheater.

The second point is that it should be ensured that the steam is completely “used” in the heat exchanger before leaving it. The complete usage of steam means complete conversion of the steam to condensate. We have to make sure there is no amount of steam remaining in the stream exiting a heat exchanger. This can be done by a placing steam trap on outlet side of the hot fluid of the heat exchanger.

Using steam as a heat transfer medium is economically justifiable when the required temperature of the heat transfer medium is less than 150 °C. When the required

temperature for the heat transfer media goes beyond this number, using steam is not economical and possibly we need to use another more expensive heat transfer medium, like non-water-based heat transfer media. These heat transfer media could be a synthetic heat transfer material like Dowtherm®.

11.5 Heat Exchangers: General Naming

Heat exchangers can be named based on different criteria.

Heat exchangers can also be named based on their types. The most common heat exchanger types are shell and tube heat exchangers, double pipe heat exchangers, plate heat exchangers, and spiral heat exchangers.

In a plant heat exchangers can be named based on the service they are working on. A heat exchanger could be a “gasoline heat exchanger” or a “treated water heat exchanger,” etc. It is very common to see the names rough heat exchanger and trim heat exchangers. Generally speaking where there are two heat exchangers in series and the first one is a process heat exchanger (as will be discussed in Section 11.8.1) and the second one is a utility heat exchanger, the upstream heat exchanger is named a rough heat exchanger and the downstream heat exchanger is named a trim heat exchanger. The reason is that the first heat exchanger brings the temperature close to the target temperature but not very close to the acceptable temperature. The second heat exchanger or trim heat exchanger brings the temperature within the acceptable range.

Heat exchangers can be named based on their functions. They can be named any of these four: cooler, heater, condenser, or vaporizer.

Heat exchangers can also be named based on their phase of the service fluids. They could be named liquid/liquid (L/L), liquid/gas (L/G), gas/gas (G/G).

A heat exchanger can be named based on the type of energy source or energy sink. They can be named utility heat exchanger or process heat exchanger. A utility heat exchanger is a heat exchanger that uses a utility stream as a heat exchange media. The utility stream could be steam, hot/cold, cooling water, etc. However, in a process heat exchanger both streams are process streams. A process stream means any stream that is not a utility stream and is one of the streams from the process.

For example, a heat exchanger in a plant could be a natural gas heat exchanger, which is an S&T HX, a heater, and a gas/gas heat exchanger, and it could be utility heat exchanger because the heating media is utility steam.

11.6 Heat Exchanger Identifiers

As it was stated in Chapter 4, the identifiers of heat exchangers on P&IDs are heat exchanger symbols, heat exchanger tags, and heat exchanger call-outs.

11.6.1 Heat Exchanger Symbol

It has been seen that in some companies they mistakenly use the PFD symbol of heat exchangers instead of P&ID symbols (Table 11.5).

In PFDs the symbol for heat exchangers doesn’t show the type of the heat exchanger. This is acceptable as in the preliminary stage of a project the type of heat exchanger is not decided and the general symbol is used in PDF. However at the P&ID stage a more specific symbol should be used.

Table 11.6 shows symbols of different heat exchangers.

There is always confusion in recognizing and differentiating between different symbols of S&T HXs. That is the reason that plenty of companies recommend referring to the heat exchanger call-out (rather than the symbol) to recognize the exact type of S&T HX.

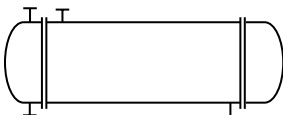
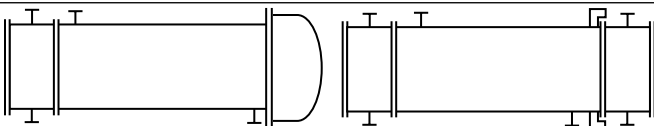
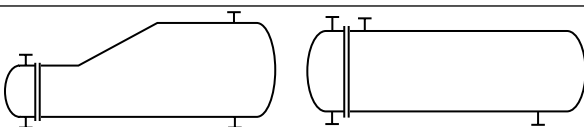
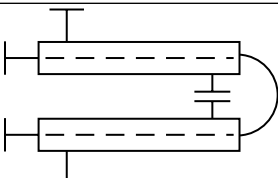
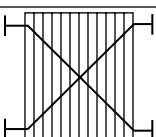
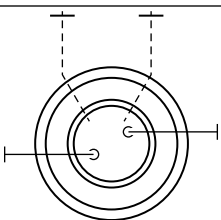
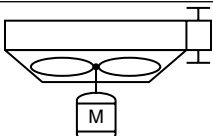
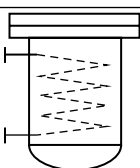
11.6.2 Heat Exchanger Tag

Not all companies use heat exchanger tags. If heat exchanger tags are needed to be shown on the body of the P&IDs, they are generally placed inside of the heat exchanger unless there is not enough room or it causes confusion.

Table 11.5 PFD symbols for heat exchanger types.

	Heat exchanger type	Heat exchanger symbol
Utility heat exchangers	Heater	
	Vaporizer	
	Cooler	
	Condenser	
Process heat exchanger		

Table 11.6 P&ID symbols for heat exchanger types.

Heat exchanger type		Heat exchanger symbol
Shell and tube (S&T) heat exchanger	Fixed head	
	Floating head	
	U-tube	
Double pipe heat exchanger		
Plate and frame (P&F) heat exchanger		
Spiral heat exchanger		
Aerial cooler		
Helical coil		

As it was mentioned in Chapter 4, the heat exchanger tag could be “E-142” or “HX- 430.”

11.6.3 Heat Exchanger Call-Out

A typical call-out for heat exchangers is seen in Figure 11.3.

- The first line is the tag number of the heat exchanger.
- The second line is the name of the heat exchanger.

300-E-143
Hydrocracker feed cooler
 Surface area: 2,700 m²
 Heat duty: 13,600 KW
 Design shell: 2,800 kPag
 Design tube: 2,000 kPag
 MDMT shell: –45°C
 MDMT tube: –45°C
 Insulation: HC/64 mm
 Trim shell: 3P8499-EAB
 Trim tube: 3P8500-CLK

Figure 11.3 A heat exchange call-out.

- The third and fourth lines are the main characteristics of the heat exchanger, which are the heat transferred in the heat exchanger (heat duty) and the heat transfer area.

Heat exchangers are mainly specified by their “heat duty,” which is the specific amount of heat transferred in the heat exchanger. Examples of heat duties of heat exchangers are 3.5 MW (megawatts), 400 kW (kilowatts), or 100 BTU h⁻¹.

A heat exchanger may also be specified by its “heat transfer area.” The heat transfer area of a heat exchanger is the total area of the “metallic enclosure” inside the heat exchanger. This area is in contact with cold fluid on one side and hot fluid on the other side, and the heat flows through it.

A specific heat exchanger may have a heat duty of 400 kW and a heat transfer area of 500 m².

- Next we have design pressure and temperature, MDMT (minimum design metal temperature), insulation, and trim. However, since a heat exchanger is basically two vessels placed inside each other (or side by side), each of the parameters mentioned should be listed for both of them.

In some shell and tube heat exchanger call-outs the TEMA (Tubular Exchanger Manufacturers Association), type of is also mentioned. The TEMA types of heat exchangers are a three letter acronym that specifies the type of head, the type of body, and the type of end of the shell and tube heat exchanger.

11.7 Heat Exchanger P&ID

In developing a P&ID it is very important to draw the P&ID as similar to the real plant as possible. For heat exchangers, in the majority of cases the hot stream goes to the heat exchanger from the top and comes out from the bottom, and the cold stream goes to the heat exchanger from the bottom and comes out from the top of the heat exchanger. We have to follow this when developing and drawing the P&ID (Figure 11.4).

The reason for this is when a stream gets colder it will be more dense and naturally it wants to go down,

therefore it totally makes sense to bring a hot stream to the heat exchanger from the top and remove it from the heat exchanger from the bottom. For the cold stream, when it gets hot it will expand and then be lighter. Therefore to follow the natural tendency of the flow it is better to bring the cold stream from the bottom of heat exchanger and remove it from the heat exchanger from the top.

This rule may have some exceptions but there are no exceptions for condensers and evaporators.

11.7.1 Vents and Drains

Heat exchangers need vents and drains. However, as a heat exchanger is basically two convoluted containers, there is a need for vents and drains on each of the enclosures. For example, in S&T HXs we need vents and drains on the shell site and vents and drains on the tube sites. As another example we need vents and drains for the cold side of a plate heat exchanger and also on the hot side of the plate heat exchanger.

Vents are valves that are placed on the top of heat exchangers and drains are valves that should be placed on the bottom of heat exchangers. Therefore a shell and tube heat exchanger could have one vent valve on the top and on the shell side and the other vent valve on the top and on the tube side and one drain valve on the bottom on the tube side and another drain valve on the bottom on the shell side (Figure 11.5).

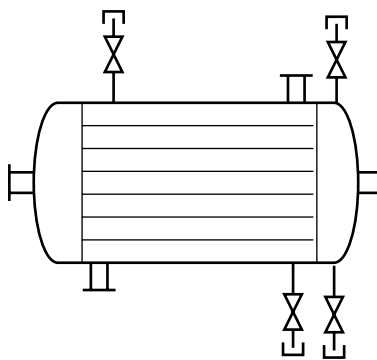


Figure 11.5 Vents and drains on a shell and tube heat exchanger.

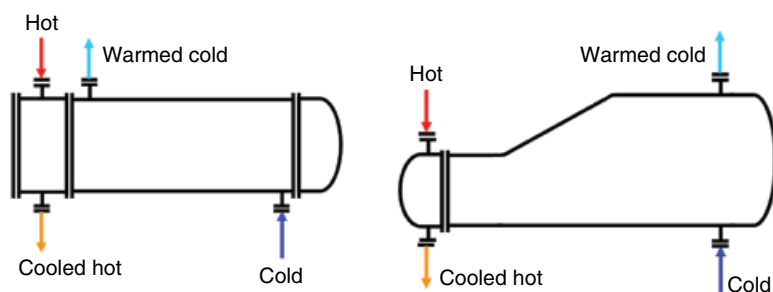


Figure 11.4 Stream arrangements in a heat exchanger.

11.7.2 Isolation Valves

The other heat exchanger attachment could be an isolation valve. Isolation valves, however, don't exist for all heat exchangers. The requirement for placing an isolation valve for heat exchangers depends on the service of the heat exchanger and the maintenance policy of the plant. Some companies don't put any isolation valves around their heat exchangers. Their logic is that: "we can't afford to pull the heat exchangers out of operation when the rest of the plant is operating thus it doesn't make sense for us to put isolation valves." This logic is not very unusual because in the majority of cases a heat exchanger cannot be shut down and streams that used to go inside of the heat exchanger bypass the heat exchanger and are sent to the next equipment without temperature change. Therefore, placing isolation valves around heat exchangers is subjective. The other issue is whether we really need to put bypass pipes around a heat exchanger or not. With the same logic the majority of companies don't put bypass pipes around heat exchangers because they cannot afford to send on a stream without its temperature adjusted. One case in which placing isolation valves and a bypass pipe around a heat exchanger is justifiable is the case where there is an available spare heat exchanger or the heat exchanger is equipped with a clean-in-place (CIP) system or a mechanical-in-place system (MIP). If a heat exchanger is equipped with a CIP or MIP system, and the adjusted temperature stream goes to a huge bulk of fluid downstream of the heat exchanger, the short period of chemical or mechanical cleaning on the heat exchanger can be affordable.

11.7.3 Chemical Cleaning Valves

If in-line and in-place cleaning is considered as part of heat exchanger care, chemical cleaning valves should be provided. Chemical cleaning valves are 2" or larger in size and can be provided as part of a heat exchanger by the vendor, or should be provided during the P&ID development stage. The chemical cleaning valves are needed for every process fluid nozzle of the heat exchanger (Figure 11.6).

The implementation of chemical cleaning valves may jeopardize the concept of the isolation of heat exchangers. Their installation should be taken with caution and after getting the client's permission.

11.7.4 PSDs

Again, as heat exchangers consist of two enclosures, both of them should be equipped with a pressure safety device to protect them from high pressure situations. For example,

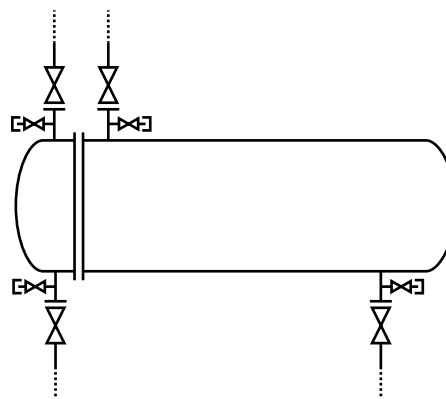


Figure 11.6 Chemical cleaning valves on a heat exchanger.

one PSD should be placed on the S&T HX and another PSV on the tube side. The PSDs are not necessarily installed on the shell side but rather they could be on the connected pipe to the shell side. For tube side PSD there is no way to install the PSD directly on the tube side therefore the PSD is installed on the connected pipe to the tube side of the heat exchanger.

11.8 Heat Exchanger Arrangement

The arrangement of heat exchangers in a process plant can be rigorously decided by pinch analysis. This concept is beyond the scope of this book. However, we talk about two common arrangements of heat exchangers.

Heat exchangers can be installed in parallel or series arrangements.

As each heat exchanger has at least two streams in it, the concepts of series and parallel heat exchangers are a bit complicated. When the concept of series and parallel heat exchangers are defined, they can be regarded one stream or two streams.

Table 11.7 shows such different arrangements. In this figure PFD symbology is used for the sake of simplicity.

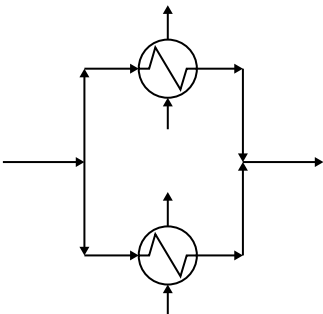
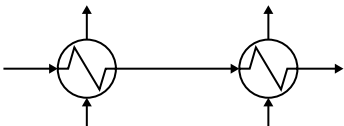
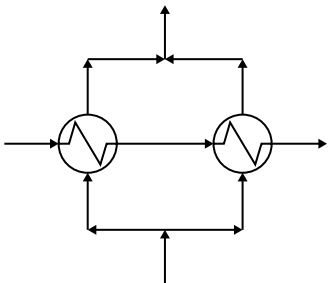
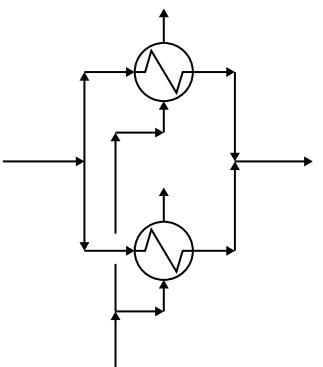
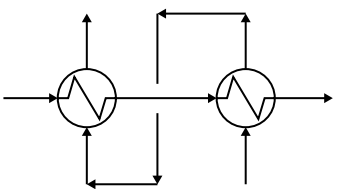
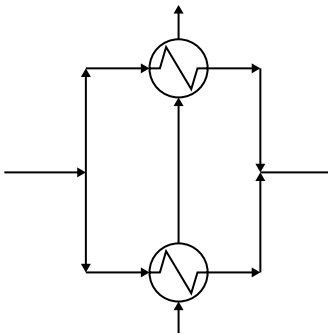
The mixed arrangements are raised here only as a deference. They are generally not used because of inefficient heat transfer in one heat exchanger.

11.8.1 Heat Exchangers in Series

Series heat exchangers could be used for different reasons.

One reason is when the temperature needs to be changed to a target temperature and the final temperature should be precisely close to the target temperature

Table 11.7 Different arrangements of heat exchangers in series and in parallel.

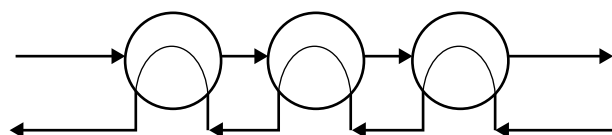
	Parallel	Series	Mixed arrangement
Regarding one stream	 Partial parallel arrangement	 Partial series arrangement	
Regarding both streams	 Full parallel arrangement	 Full series arrangement	

selected by the designer. In such cases a “rough temperature adjustment–trim temperature adjustment” could do the job. The rough temperature adjustment is done in a process heat exchanger and the trim temperature adjustment is done by a utility heat exchanger in series and downstream of the process heat exchanger.

The other use of series heat exchangers is when a temperature change is needed and there are several process streams available but none of them can solely change the temperature to the target temperature. In such cases several heat exchangers may be put in series so that each of them works with one of available process streams to finally reach the target temperature.

The third use of heat exchangers in series is to provide a large temperature change in a heat exchanger. Generally speaking, when a large temperature change is needed (say more than 150°C) it cannot be handled by one heat exchanger therefore several heat exchangers could be placed in a series arrangement.

In the cases where several heat exchangers are placed in series, and they are in series regarding both streams (full series arrangement), it is very important to know that it is always the coldest stream that is brought in

**Figure 11.7** Arrangement of a series heat exchanger regarding both streams.

contact with the hottest stream and therefore the arrangement of the two streams in the series heat exchangers are against each other. This is to maximize the heat transfer of the heat exchangers. Figure 11.7 shows this concept with PFD symbology.

In some full series arrangements the heat exchangers are stacked up. This means that two or more heat exchangers are placed on top of each other in a way that a stack is created. This practice saves some of the plot plan. This could be the case mainly where the heat exchangers are the same size. The P&ID schematic of stacked heat exchangers is shown in Figure 11.8.

In such cases there is no need to put vent and drain valves on each single heat exchanger. We only need to put vent valves on the top heat exchanger and drain

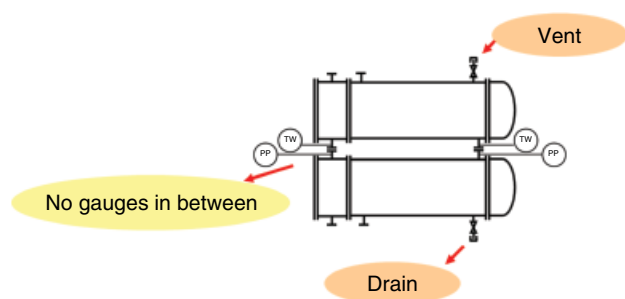


Figure 11.8 Stacked heat exchangers.

valves on the bottom heat exchanger. Generally there is not enough room to place drain or vent valves between the stacked heat exchangers. Even chemical cleaning can be reduced in this arrangement.

As there is generally not enough room between the two stacked heat exchangers, the sensors can be deleted and only the “location” for the portable sensors are left. The concept of location for portable sensors will be covered in Chapter 13.

11.8.2 Heat Exchangers in Parallel

Heat exchangers can be placed in parallel when all of them are operating, or in some less common cases to provide spare capacity for the heat exchanger.

If the flow rate to a heat exchanger is huge, and/or the required temperature change is high, a heat exchanger with an overly high heat transfer area may be needed. Each type of heat exchanger has a limitation on the heat transfer area. Therefore, sometimes splitting the flow rate and putting two or more heat exchangers in parallel may be needed to avoid placing a heat exchanger that needs an overly high heat transfer area. The limitation on the heat transfer area for each heat exchanger is dictated by technical and economical factors but sometimes can be deviated from. For example very large heat exchangers with very large heat transfer areas will require bringing a heavy duty crane to the plant when the heat exchangers need to be sent to the workshop for maintenance. Large cranes are not available in all plants and usually only large plants can afford to keep them. Therefore, we can have multiple heat exchangers in parallel and all of them are working, for example, $2 \times 50\%$ applications.

The other case where we may have parallel heat exchangers is when we need a spare heat exchanger. Generally speaking, heat exchangers are too expensive to allow us to keep them as a spare in plants. Therefore it is very rare to see a parallel heat exchanger $2 \times 100\%$, which means one operating heat exchanger and another spare heat exchanger. However, in some cases, for example if the service fluid is very fouling, we may have a spare heat exchanger.

11.9 Aerial Coolers

If it is decided to use ambient air as the cooling medium on the shell side of a S&T HX, the type of heat exchanger can be used is an “aerial cooler.”

In aerial coolers, a bundle of tubes (generally finned tubes) is secured in a frame and there is no shell at all. Air blows through the tube bundle with the help of fan(s). Aerial coolers have the advantage that they have no need for a cooling medium because they use ambient air for cooling purposes. However, they need electricity for the operation of the motor that is connected to the fan.

Aerial coolers are generally used in multiple units in plants.

The smallest component of an aerial cooler is the “tube bundle.” Each tube bundle has one set of dedicated headers, inlet header and outlet header. Each tube bundle may have one or two (even up to four) inlet process flows and the same number of outlet process flows (Figure 11.9).

A “unit” is several side-by-side tube bundles that work as a single piece of equipment and carry one tag on the P&ID.

Aerial coolers can be seen in process plants as “banks,” which are large pieces of equipment. Each bank of aerial coolers can be more than one tagged aerial cooler in a P&ID. One (or more) specific area of a plant may be dedicated to aerial coolers and all the aerial coolers of the plant will be located there as a “bank” (Figure 11.10).

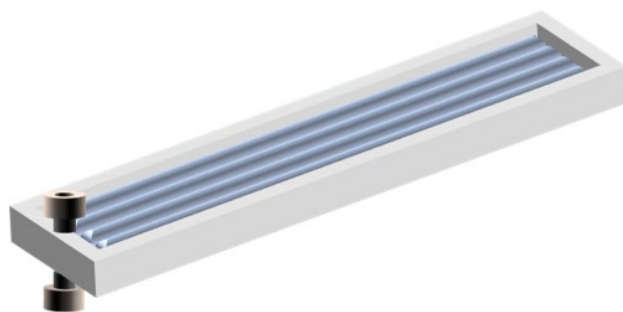


Figure 11.9 The tube bundle of an aerial cooler.

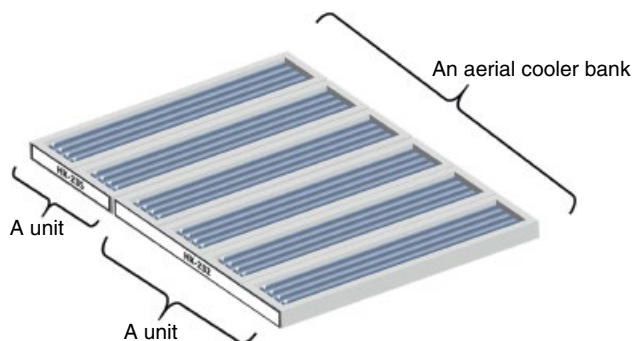


Figure 11.10 The unit and bank in aerial coolers.

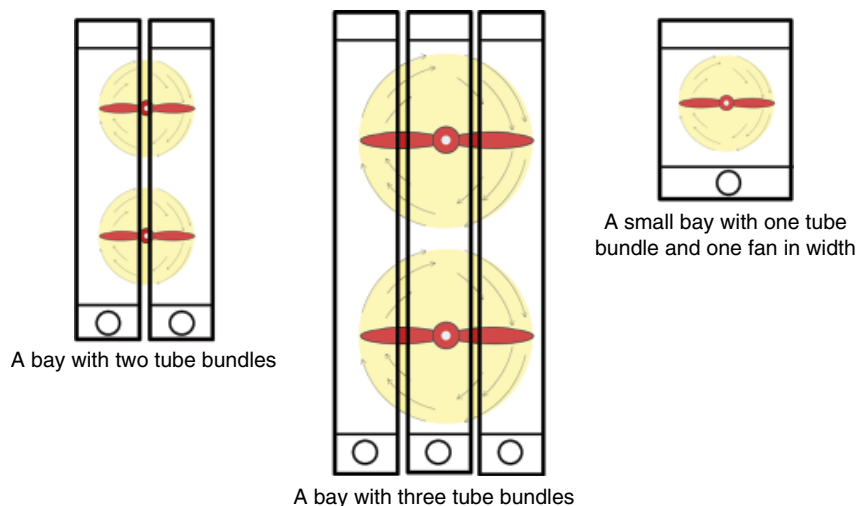


Figure 11.11 Definition of bay.

A plant however may have several aerial cooler farms.

A “bay” is more mechanical concept and means every side-by-side set of tube bundles that have one fan in their width

There is generally another fan on the length of each tube bundle in the majority of cases, except in very small aerial coolers where only one fan serves a few tube bundles (Figure 11.11).

In the P&ID each single tag of an aerial cooler is a unit and can consist of several bays and several tube bundles.

The aerial cooler banks cannot be recognized in P&IDs.

A typical call-out for heat exchangers is seen in Figure 11.12.

- The first line is the tag number of the heat exchanger.
- The second line is the name of heat exchanger.
- The third and fourth lines are the main characteristics of the heat exchanger, which are the heat transfer area and the heat transferred in the heat exchanger (heat duty).
- Next there is design pressure and temperature, MDMT (minimum metal design temperature), and trim.
- The last item is the driver power. The driver is generally an electric motor.

The number of bays may also be mentioned in the aerial cooler call-out.

22-AE-0013
 SPILLBACK COOLER
 HEAT DUTY: 7,100 kW
 BARE SURFACE AREA: m²
 DP/DT: 1300 KPag/FV @ 180°C/-29°C
 DRIVER: KW

Figure 11.12 An aerial cooler call-out.

11.9.1 Aerial Cooler P&ID

As each aerial cooler may have several tube bundles thus the process flow should be evenly distributed amongst them. Therefore flow distribution is a critical issue. To provide a good distribution – and collection – the distribution headers are placed above the bank of aerial coolers and the collection headers below the coolers are placed (Figure 11.13).

As the P&ID is a 2D diagram and is generally a plan view, it is not always easy to show the headers. In some P&IDs the distribution header and collection header are completely ignored and left on the shoulders of pipe modelers to “recognize” the existence of them. In some other P&IDs the cross section of the header is shown (Figure 11.14).

The other details of aerial coolers are drains and vents, isolation valves, and pressure safety devices (Figure 11.14).

Each tube bundle of an aerial cooler needs drain and vent capability. Sometime the aerial cooler by itself has drain and vent valves on its headers. However this is not the case for all aerial coolers. Then, the designer may

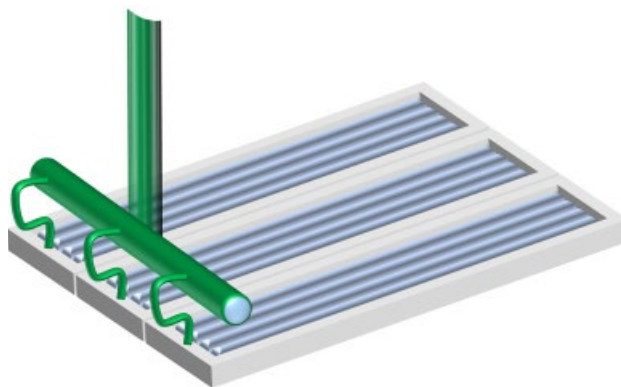


Figure 11.13 Flow distribution in aerial coolers.

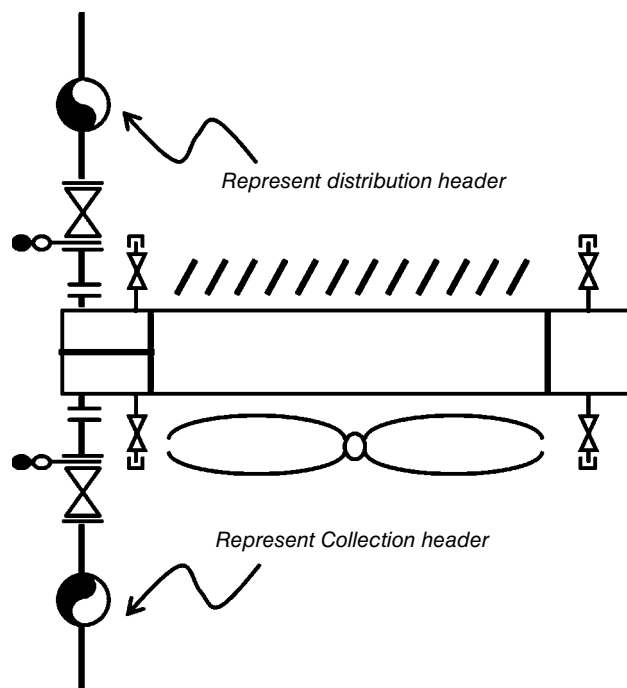


Figure 11.14 P&ID detail of an aerial cooler.

need to provide air and drain valves on the connecting pipes of an aerial cooler. If this is the decision it is appropriate to provide one set of vent and drain valves on each header of a tube bundle. This means that each tube bundle has two vent valves on the top of each header and two drain valves on the bottom of each header.

An aerial cooler can also be equipped with isolation valves.

As the tube bundle of aerial coolers is considered as an enclosure, it should be protected from over-pressurization by installing PSDs.

11.9.2 Dealing with Extreme Temperatures

As aerial coolers use ambient air generally without any pre-treatment, they thus should be robust. There are, however, some extreme cases that should be considered in the design of aerial coolers and/or in the P&ID development of them. The main important issue is extreme ambient air temperatures.

The ambient temperature change depends on seasons and going from night to day or vice versa. The control system around an aerial cooler tries to compensate the ambient air temperature so that there is a fairly fixed temperature of process fluid exiting the aerial cooler. Such a control system will be discussed in Chapter 13. Here, however, we talk about methods to deal with “extreme” ambient temperatures; how handle the cases where the ambient air is extremely high or extremely low,

which may causes frosting if there is enough vapor in the air. If such conditions are forecastable in the area, it should be considered in the design.

- Dealing with extremely low temperatures: extremely low temperatures (say less than $15-20^{\circ}\text{C}$) could be problematic, mainly where there is a chance of frosting on the aerial cooler tubes. There are mainly two ways to handle such conditions: using a steam coil and warm air recirculation.

In the steam coil solution, a bundle of steam coils is superimposed on the aerial cooler bundles. By doing this, the air moving through the aerial cooler bundle has to go through the steam coils at the beginning. This warms up the cooling air and prevents frosting.

In the warm air recirculation solution, instead of “virgin” ambient air, the warmed ambient air is used for cooling purposes in the aerial cooler.

There could be two sources of warmed up air. In “internal air recirculation” the source of warmed up air is the already used ambient air after some cooling effect on the aerial cooler. In “external air recirculation” the source of warmed up air is partially from the “air conditioned air” of the adjacent building. Obviously the external air recirculation is available only for aerial coolers in the vicinity of air-conditioned buildings.

Implementing the concept of air recirculation needs putting several louvered walls around the aerial cooler in a way that it looks like the aerial collar is placed inside of a “box” with perforated walls.

One example of a boxed aerial cooler is shown in Figure 11.15.

A combination of steam coil and recirculation air can also be used.

In the P&ID, the louvers should be shown and tagged in a similar way to control valves.

- Dealing with extremely high temperatures: I have seen in extremely hot days of summer that water hoses have been placed on the aerial cooler bundles! However a factory-built system to deal with extremely high temperature is a “humidifier aerial cooler.”

11.9.3 Aerial Cooler Arrangement

Aerial coolers in series are generally avoided. This is because of difficulties in the symmetrical distribution and collection network between two aerial coolers in series considering the limited available space. However it is not difficult to avoid aerial coolers in series. By using several aerial coolers in parallel (which is very easy) and also using a multiple pass configuration, series aerial coolers can easily be avoided.

As aerial coolers have a modular structure, in the majority of cases they are installed in a parallel arrangement.

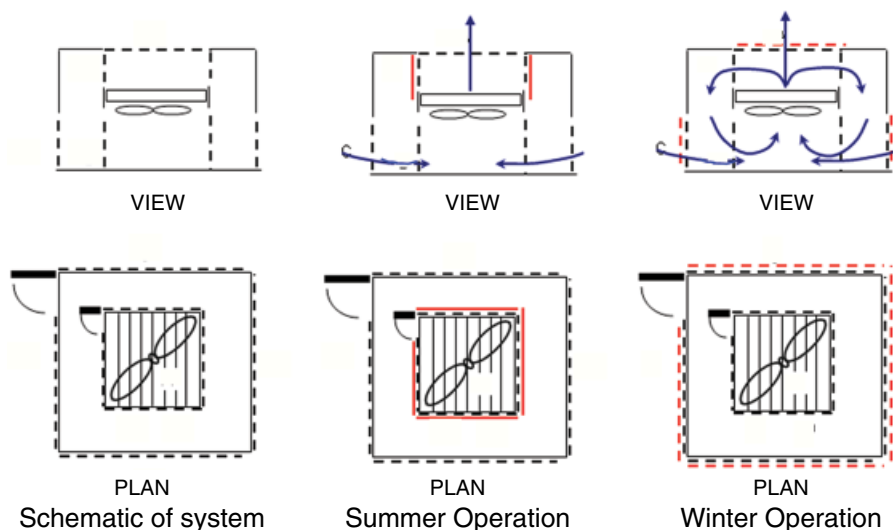


Figure 11.15 Aerial cooler with the air recirculation concept implemented.

There are, however, applications where an aerial cooler is used for very small heat duties, which need only a single bay aerial cooler.

11.10 Merging Heat Exchangers

Even though heat exchangers are expensive pieces of equipment they are not generally merged. The reason is that the design of heat exchangers is so specific that it is not easy to find one heat exchanger that matches two different services.

11.11 Wrapping-up: Addressing the Requirements of a Heat Exchanger During its Life Span

In this section we check our design to make sure we have covered all the needs of heat exchangers during each phase of plant life. As was discussed before, these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item.

- 1) Normal operation of heat exchangers: the required considerations are already covered.
- 2) Non-routine operation (reduced capacity operation, start-up operation, upset operation, planned shut-down, emergency shut down): these phases of plant needs two components to be handled: the process component and the control/instrumentation component. The process component is discussed here but the control/instrumentation component will be discussed in Chapters 13–15.

One main issue regarding heat exchangers is that when flow goes below a specific value, the heat transfer is decreased. If there are heat exchangers in parallel it may be decided to pull one unit out of operation to keep the adequate flow rate in the second parallel heat exchanger. However, if there is no parallel unit, it is not easy to mitigate the problem of a very low flow rate.

- 3) Inspection and maintenance: general consideration regarding inspection and maintenance of all items will be covered in Chapter 8. Here, however, we cover the specific requirements.

There could be a piping arrangement to allow “back-flush” of a heat exchanger during normal operation. This back-flush arrangement can be placed on cooling water to remove fouling from the water side of the cooler (Figures 11.16).

- 4) Running the plant in the absence of pipe: There is generally no consideration for the times that a heat exchanger is out of operation. The reason is that heat

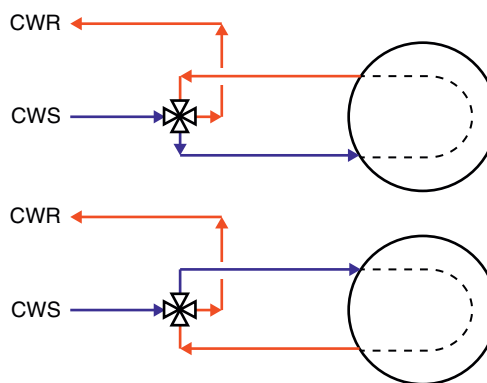


Figure 11.16 Back-flushing in heat exchangers.

exchangers are expensive pieces of equipment. When a heat exchanger is out of operation, it generally impacts the operation.

11.12 Fired Heaters and Furnaces

Furnaces are the least favorite heat transfer units for different reasons; they are expensive and they have a bunch of safety issues to be addressed. They may have names other than furnace; when the process fluid is water, they are named boilers or steam generators.

A fired heater is a heat transfer unit that uses fire to heat up the temperature of a process fluid.

If a fired heater is intended to be used to change the process fluid physically or chemically to other fluids by heating up, it can be named a furnace.

The main component of every furnace is the heating system. There are mainly two types of furnaces: fired heaters where the heating system is fire, and electric heaters where the heating system is an electrical element. Electrical furnaces are expensive from a capital cost and operating cost viewpoint and are not used unless a very high temperature is needed. Therefore our discussion is limited to fire heaters.

Fire heaters use fire to increase the temperature of process fluids. Fire is generated in burners where fuels and oxidant are brought together. The fuel could be anything from oil, gas, or coal to some specific sludge. The oxidant is mainly air and sometimes other oxidants like oxygen.

The result of burning fuel is flue gas. There are some furnaces where their source of energy is not burning fuel but electrical energy.

Therefore, a furnace has at least three systems that should be taken care of during the P&ID development: the process fluid side, the firing side, and the flue gas side (Figure 11.17).

11.12.1 Process Fluid Side

As it was mentioned, for efficient heating, the fluid should be exposed to heat with narrow thickness.

However, there are some cases where a huge stream needs to be heated in a furnace. In such cases the stream is split into several narrow bore pipes before entering the furnace. This multi-pass arrangement is needed to make sure the coils inside the fired heater have a small enough diameter (generally less than 4–6 in.) to make sure heat is absorbed effectively by the process fluid. Therefore each

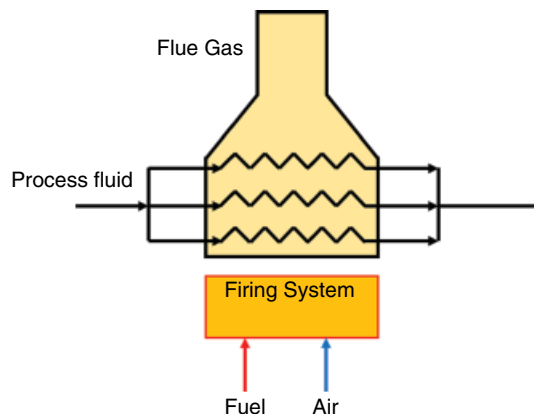


Figure 11.17 Three fluid streams within a fired heater.

fired heater may warm up a stream, not through a single coil, but rather through several narrow bore coils of generally less than 4–6 in. Each coil circuit in a furnace is named one “pass.” We may have a furnace with 2, 4, 6, etc. passes. Therefore, a large-bore pipe for a process fluid is split into 2–6 (or more) small-bore pipes outside the fired heater, and then all of them enter the fired heater chamber. These multiple passes will merge together again on the other side of the fired heater after heating the process fluid.

One responsibility of the P&ID developer is to distribute the stream of interest for heating amongst several coils evenly. This can be done by a control system or by symmetrical piping. These solutions are briefly discussed in Chapter 6 and the control system is discussed in more depth in Chapter 15.

Furnaces are designed and built to handle large thermal expansion of its elements because of large temperature changes. However furnaces cannot handle the thermal shock. Introducing cold process fluid into the furnace coils may cause thermal shock and initiation of cracks and rupture in the furnace coils. Because of that when the temperature difference between the cold process fluid and the furnace target temperature is more than 100°–150°C, preheating may be needed. This preheating could be necessary to bring the cold process fluid to a temperature closer to the furnace temperature. The preheating is generally done by a heat exchanger and more commonly this heat exchanger uses the hot process fluid exiting the furnace to heat up the cold process fluid entering the furnace. Sometimes preheating is done through heat exchange from the hot flue gas of the furnace.

11.12.2 Flue Gas Side

Flue gas is generated inside the furnace and directed to the atmosphere through the stack.

Flue gas generally finds its way out of the furnace box by going up. Flue gas is hot and an adequately long stack provides enough draft (natural draft) to suck the flue gas out of the furnace box. The other option of removing flue gas from the fired heater box is placing a fan to suck the flue gas to the outside or an “induced draft” system.

As the flue gas is hot, releasing it to the atmosphere means losing energy. Thus, it is not rare to see another set of coils in the flue gas route toward the stack to use – at least a portion of – the flue gas energy.

One common problem of coils in the flue gas route is soot accumulation. Accumulation of soot around the external wall of the flue gas coils decreases the heat transfer and also may cause overheating in some spots of the coils. Soot blowers are briefly discussed in Chapter 8. A soot blowing system is placed to remove the accumulated soot at specific intervals. Soot blowers work intermittently based on a pre-determined schedule. Their operation happens during the normal operation of a furnace and without major interfere. Soot blowers generally spray steam toward the coked coils to remove the soot.

Generally things in the route of the flue gas are installed on furnaces and we don’t need to add anything on the P&ID.

11.12.3 Firing Side

Having fire in the process element looks a very scary thing. That is the reason why there are bunch of provisions to make sure the fire is under control at all times. Some of these provisions are provided by a safety instrumented system (SIS) and some others are provided by good design. The SIS is discussed in Chapter 15.

The firing side comprises at least two separate systems: the fuel side and oxidant side.

Fuel is brought to the burners through a fuel distribution system. Two very common fuels for furnaces are fuel oil and fuel gas. If the fuel is coal, it should be brought near the furnace, not through the utility network, but through other methods, which are beyond the scope of this book.

If the oxidant is air, it could be brought to the burner by natural draft, forced draft, or only relying on the induced drafting system of the flue gas.

The air could be pre-heated through coils in the convection section of the fired heater for more efficient burning.

The firing happens in burners. As fuels are commonly fuel gas or fuel oil there are fuel oil burners and fuel gas burners. There are also burners that are designed to use either of those fuels and thus are “dual fuel burners.”

Because in microscopic scale fire happens in the vapor phase (rather than the liquid phase), the fuel oil – and not the fuel gas – should be converted to vapor or something else easily convertible to vapor. As converting fuel oil to vapor is a very energy intensive operation we prefer to convert them to small droplets, which can then be easily

evaporated in the temperature provided by firing. Therefore, all fuel oil burners need a type of “atomization.” From a theoretical viewpoint, to convert a continuous fluid to discrete droplets you need to exert energy in it. The energy for atomization can come from a high pressure stream or by a mechanical device. Amongst high pressure fluid atomization systems we have pressure jet atomization (with pumped fuel oil), stream atomization, and air atomization. The whole purpose of a discussion on atomization here is to point out that sometimes the P&ID developer needs to bring another stream (other than fuel) to the burner for the purpose of atomization.

Burners need a “pilot flame” to help the firing as soon as it is ordered to do so. If you go down deep to the theory of fire you will see that starting a fire depends heavily on probabilistic phenomena. This means when everything is ready to start a fire (which means the fire triangle is available), it may take a “short time” to start firing. This short time could be acceptable when you try to fire your barbeque but in large furnaces, releasing a fuel and air mixture without fire is dangerous. Therefore each burner is equipped with a small flame near the main burner tip, which is fueled by an easy-to-get-fire fuel (generally fuel gas) separate from the main fuel pipe. This small built-in burner is named a “pilot.” Therefore during P&ID development fuel gas is also needed for the pilots from a separate network.

As mentioned before the fuel to burners are provided by fuel gas or fuel oil utility distribution networks. Figure 11.18 shows a typical summarized P&ID of the fuel route to a burner.

Isolation valves should be provided right after the fuel header, before the burner header and upstream of each burner.

Small valves should be provided (at least two) for draining, venting, and or purging the pipe.

The fuel preparedness block could be different things depending on the type and condition of the fuel. It could consist of a pressure regulator to drop the pressure to a fairly constant pressure. A fuel filter could also be here if the fuel could have particles that could plug the burner tips.

Blocks of switching valves are provided on the fuel sub-header and on each burner branch. In some designs two switching valves are used in series to satisfy the required safety level. In some older designs there was no additional set of switching valves on each burner branch.

A block consisting of a control valve station, is provided on the pipe in series with the switching valve block.

The pilot gas route doesn’t need a control valve station but it still needs a switching valve.

It is very important to mention that in plenty of jurisdictions BMS (burner management system) is a code and therefore firing systems are highly regulated.

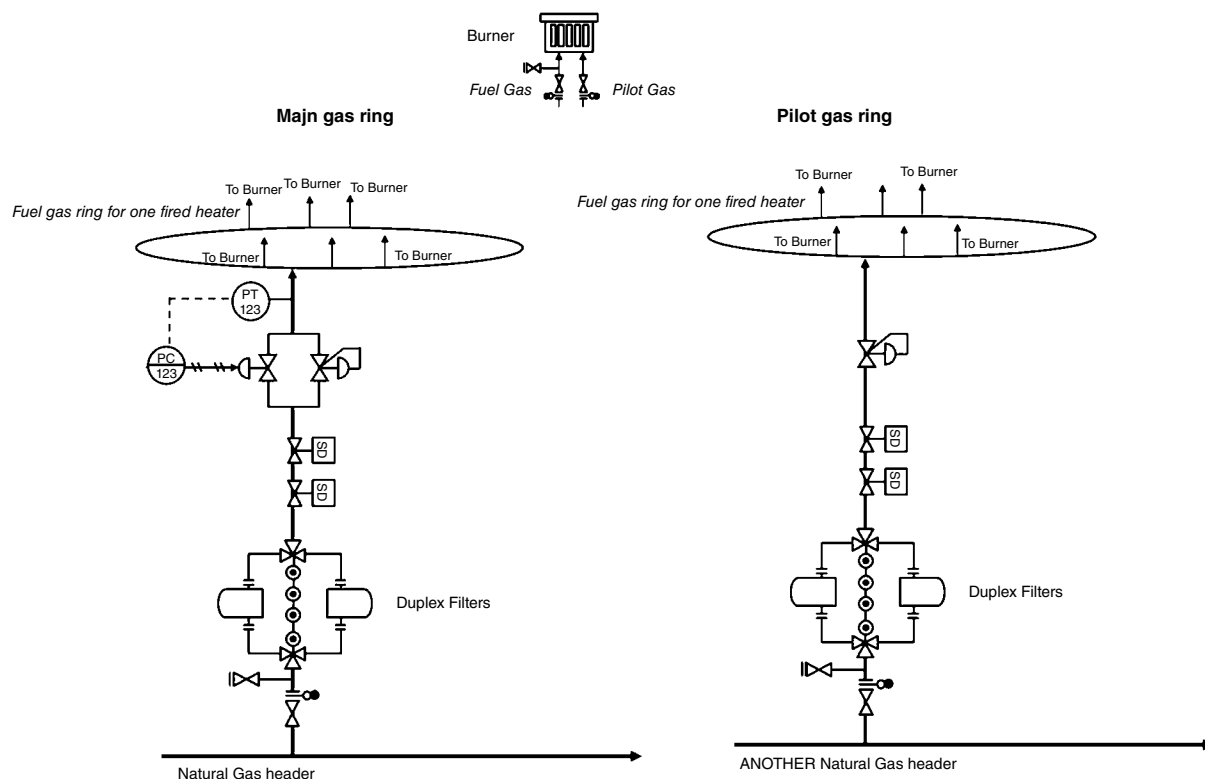


Figure 11.18 P&ID of a fuel gas burner with pilot gas.

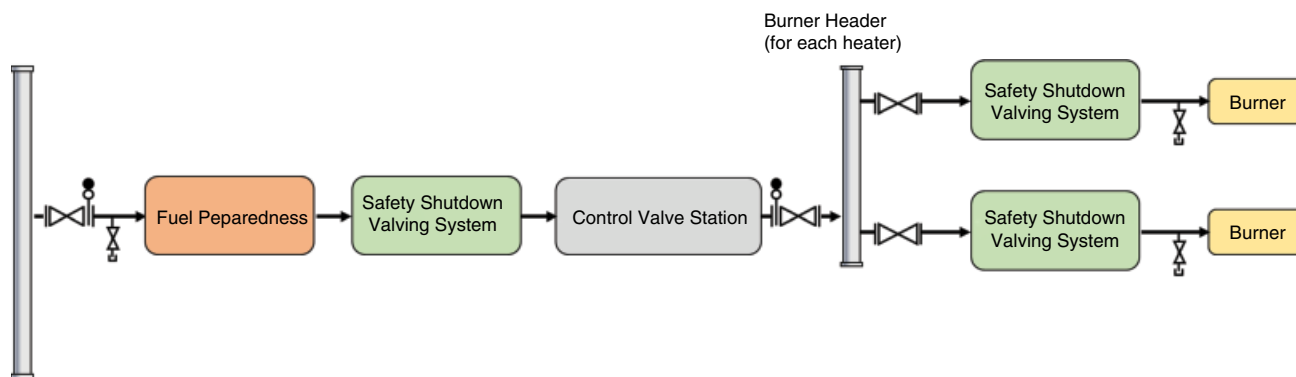


Figure 11.19 Fundamentals of the fuel route to burner.

Generally speaking there is more than one burner in each fired heater, therefore a burner header may be needed. It is very important to provide the fuel to several burners of a fired heater with the same flow and pressure. This required distribution is specific to the fuel pipe of burners.

It discussed in Chapter 17, there are two main types of flow distribution, tree type and loop type. As there is more chance of pressure and flow swinging in fuel oil – rather than fuel gas – the fuel oil is distributed amongst the several burners of a fired heater through a loop distribution. A ring header around the furnace provides a fairly even fuel oil for the burners.

Figure 11.19 shows the P&ID of a fuel gas burner with pilot gas.

As part of SIS, if for whatever reason the flame put off, the unburnt fuel-air mixture inside of furnace should be displaced as soon as possible. The steam can be used for this purpose. This steam is known as “snuffing steam.”

11.13 Fire Heater Arrangement

Fired heaters are rarely placed in series or parallel. Fired heaters are such expensive pieces of equipment that it is preferred to only have one of them in service.

However there is a very common case in which a heat exchanger and a fired heater are in series.

The logic for heat exchangers–fired heaters in series is the same as for heat exchangers– heat exchangers in series mentioned before; if a large temperature jump is needed (say more than 150°C), heat exchangers–fired heaters in series may be needed to prevent any thermal shock in the fired heater.

11.14 Merging Fired Heaters

Furnaces are generally not merged together because each of them is tailor-made for the requirements of each stream.

11.15 Wrapping-up: Addressing the Requirements of Fired Heaters During their Lifespan

In in this section we check our design to make sure we cover all the needs of containers during the defend phases of the plant. As was discussed before, these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item.

The requirements are checked here:

- 1) Normal operation: the required considerations are already covered. One important item for fired heaters

is “peep holes”, which provide a means for the field operator to check the flame.

- 2) Non-routine operation (reduced capacity operation, start-up operation, upset operation, planned shut-down, emergency shut down): these phases of the plant need two components to be handled: the process component and the control/instrumentation component. The process component is discussed here but the control/instrumentation component is discussed in Chapters 13–15.

Because there is fire inside of fired heaters there are plenty of safety measures around them. Although the majority of them are already implemented by the vendor, it is important to study each of them before implementing them in the P&ID to make sure they follow the codes and/or client safety rules.

- 3) Inspection and maintenance: general consideration regarding inspection and maintenance of all items is covered in Chapter 8.

There are plenty of activities regarding fired heater and boiler inspection and maintenance. One main activity is regarding the integrity of the refractory layer inside of the firing box.

- 4) Operability in the absence of one item: there is generally no spare provided for fire heaters as they are very expensive process items. If for whatever reason a tank goes out of operation the unit may need to be shut down.

12

Pressure Relief Devices

12.1 Introduction

In Chapter 7, different safety related valves were discussed. Here we talk about the detail of pressure safety valves.

12.2 Why Pressure Is So Important?

As mentioned in Chapter 5, in each plant for each single piece of equipment there could be five process parameters that are swinging. For each parameter there are guards to limit the extent of swing and to push back the parameter to its normal level. Placing mechanical relief is the last guard before the occurrence of a disaster (Figure 12.1).

The process parameters are flow, level, pressure, temperature, and composition. As the most common mechanical relief is the “pressure safety valve” you may ask what is more important about pressure than the other process parameters (Figure 12.2).

The point is that changing every single process parameter will change the pressure.

Increasing the flow in a pipe will increase the pipe pressure.

Increasing the liquid level in a container will fill the container and eventually increase the pressure of the container.

The increasing temperature may be twofold. On the one hand, by increasing the temperature generally the fluids expand and they will increase the pressure. On the other hand, the higher temperature generally weakens the body wall of the process element enclosing the fluid and makes the enclosure easier to burst.

Changing the composition not always, not quickly, and not directly affects the occurrence of a high pressure disaster. One example of the effect of composition change on high pressure disaster is when the composition of a non-acid resistant vessel suddenly turns into an acid. In this case the vessel wall becomes corroded and gets thinner and ready to burst.

12.3 Dealing with Abnormal Pressures

Abnormal pressure here means any over-pressure or under-pressure conditions happening for enclosures.

Generally speaking, to deal with any issue there are four different solutions available: (i) inherent solution, (ii) passive solution, (iii) active solution, and (iv) procedural solution.

For example, if the problem is “infectious diseases,” there are a few solutions available:

- Inherent solution: relocating to another planet with no such diseases
- Passive solution: increasing the strength of your body’s immune system
- Active solution: taking medicine when you get sick
- Procedural solution: follow this procedure: “don’t go to tropical areas”.

Similarly, there are four different groups of solutions available when dealing with over- or under-pressure situations. Each of them have their own set of costs and benefits, and some are more suitable than others for various situations.

- Inherent solutions: An inherent solution is simply the elimination of the problem, or the conditions that cause the problem. In the case of removing overpressure, an inherent solution is “either not working at high pressure at all, or not using any enclosed system.”

The inherent solution is often not generally within the realms of possibility. However, this solution can be used to decrease the extent of the issue, by using a low-pressure process rather than a high-pressure one. Such a change could be costly and may only be implementable in the very early stages of a project.

- Passive solutions: passive solutions typically take the form of fortifying the affected elements before they are exposed to the problem. In a plant, a passive solution for high pressure is to design all the equipment to handle the maximum amount of pressure that could

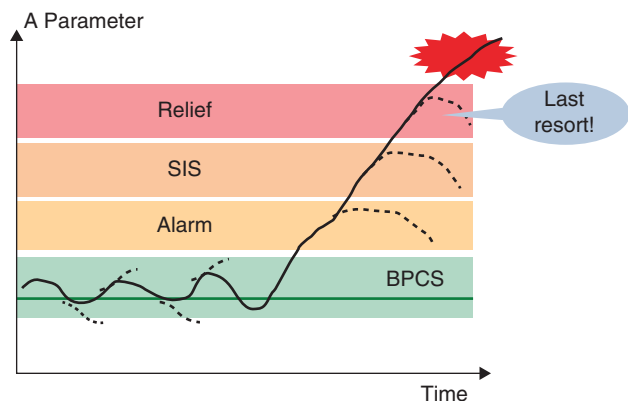


Figure 12.1 Process parameter “guards.”

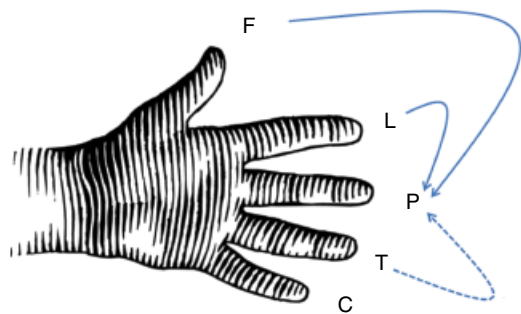


Figure 12.2 Importance of pressure as the “index” for other process parameters.

possibly be exerted by the process, or all “over-presurizing scenarios.” This means that all the equipment in a plant would have very high “design pressures” or very thick walls.

- Active solutions: active solutions are types of actions that work against the main cause of the problem. In this context, active solutions are the set of solutions previously available in process plants: basic process control systems, alarm and operator intervention, safety instrumentation systems, and of course, pressure relief devices.
- Procedural solutions: in a procedural solution, a safety instruction is written and communicated to the relevant operators. This is the weakest way of dealing with a problem. If a high-pressure case is handled via this

solution, then “operational procedures” need to be established and issued to the plant operators for each system, stating which valves should be opened in the case of high pressure. This solution is absolutely unacceptable in dealing with high pressure because of the high level of monitoring required, and the very quick action required by operators.

As may be clear by now, the most viable solutions for dealing with overpressure or under-pressure situations are passive and active solutions. In this context, implementing the passive solution means designing all elements of a plant with a design pressure that is higher than all predictable overpressure scenarios, whereas implementing the active solution means installing pressure relief devices on all equipment and elements that are potentially exposed to high pressure. Each of these two solutions has its pros and cons.

The biggest advantage of a passive solution is peace of mind; if a plant is designed based on the maximum attainable pressure, there would be no reason for concern. However, this has some drawbacks. First of all, it is not always possible to predict the highest attainable pressure in different overpressure scenarios. Second, such a plant would be very expensive, because of the thick walls required for all of its elements. Finally, the selected wall thickness needs to be continuously checked to make sure that there is no loss in wall thickness, which may reduce the nominal design pressure.

The active solution, i.e. installing pressure safety devices in different locations around the plant, is a good idea since a single PRD (pressure relief device) protects the system against all known and unknown overpressure scenarios, even without knowing the maximum attainable pressure during those scenarios. It is also good for regulatory bodies as they only need to check the integrity of a single PRD, rather than checking wall thickness, making sure no overpressure scenario is missed during wall thickness calculation, etc.

However, it also has some disadvantages in that the PRDs need to be checked periodically and their correct functioning needs to be validated. A summary of the features of passive and active solutions in dealing with high-pressure (or low-pressure) scenarios is shown in Table 12.1.

Table 12.1 Passive and active solutions.

	Pros	Cons
Passive solutions	<ul style="list-style-type: none">• Greater peace of mind due to the nature of the solutions: it’s implemented and then left alone	<ul style="list-style-type: none">• Expensive solution• Needs periodical design validation (e.g. is this container wall still standing up to the pressure?)
Active solutions	<ul style="list-style-type: none">• Mitigate all known/unknown safety issues based on a set acceptable limit• Easier for monitoring by regulatory bodies	<ul style="list-style-type: none">• Needs periodic validation of safety systems• Needs periodic recalibration

12.3.1 Active Versus Passive Solutions

The above discussion shows that using only one solution (either passive or active) to deal with high-pressure scenarios may not be a good idea; rather, a combination of these two methods should be used in order to have a safe and economical plant. Therefore, the allocation of issues to these two solutions is the primary task. A summary of the task allocation is shown in Table 12.2.

The concepts stated in Table 12.2 are expanded upon in the sections below.

12.3.2 Where Could Passive Solutions Be Used?

There are not very many cases where the passive solution, or fabricating the equipment based on the highest attainable pressure as the design pressure, is acceptable. This solution can basically be used where it is legal and where a worst-case maximum attainable pressure exists, and can be calculated. Generally speaking, the regulatory body's preference is to use active solutions, and they generally prefer to "see" a PRD on every single container. It is not easy to estimate the maximum attainable pressure for all overpressure scenarios.

For example, in a fire scenario, it is difficult to estimate the maximum attainable pressure because of the unpredictable nature of fire. The other example is a blocked outlet of positive displacement (PD) pumps. When a valve on the outlet of a PD pump is accidentally closed, the pressure on the discharge side of the pump will increase. Here there is no "maximum attainable pressure"; pressure will increase until the pump casing ruptures. In this case, the use of a passive solution is also impossible.

The other example is protecting a centrifugal pump against high pressure caused by a mistakenly closed valve on its discharge side. In this case, the maximum attainable pressure can easily be estimated and it is what we call the "pump shut-off pressure" or "pump dead-head pressure." This case could be a good case for using a passive solution to protect the piping and equipment

downstream of the centrifugal pump. To do that, the downstream piping and equipment would need to be fabricated based on a design pressure equal to the centrifugal pump dead-head pressure.

Some examples of cases where passive solutions have been used are internal fire, (external) jet fire, hydraulic hammer, and blocked outlet of centrifugal pumps.

12.3.3 Where Should Active Solutions Be Used?

The short answer to this question is: in all enclosures. The long answer would add to that: in high-pressure systems as much as possible, unless it is not technically feasible.

For example, active solutions (i.e. installing PRDs) can be used to protect against a fire pool (a fire that engulfs the equipment), thermal expansion in pipelines, control valve failed or jammed open, and blocked outlets. Typical scenarios that are unlikely to accommodate active solutions include some gas vessels and underwater scenarios, where there is no room for release.

12.4 Safety Relief System

As soon as a pressure relief device is installed on the first pressure safety valve on a process item, a "process relief device system" should be developed (Figure 12.3).

PRDs release fluids from the inside of enclosures to the outside. Then such released fluid should be collected through a collecting system and directed to a specific type of disposal system that is named an "emergency disposal system."

The collecting system is a type of pipe network and will be discussed in Chapter 16 as part of utility networks.

Emergency disposal systems are briefly discussed at the end of this chapter.

Table 12.2 Application of passive and active solutions.

Applications	
Passive solution: process design	<ul style="list-style-type: none"> ● If maximum attainable pressure exists and is specifiable ● If it is legally acceptable
Active solution: installing a PRD	<ul style="list-style-type: none"> ● As much as possible ● Unless it is not technically doable (e.g. some gas vessels, underwater requirements)

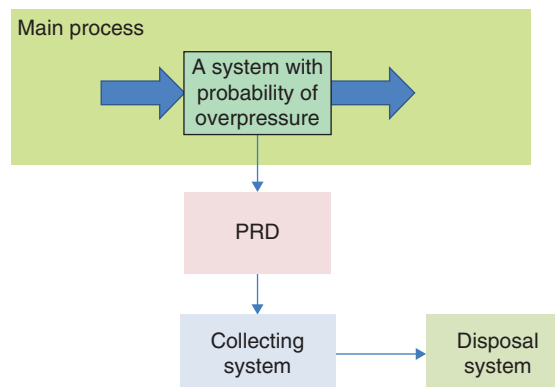


Figure 12.3 Pressure relief device system.

12.5 What Is an “Enclosure,” and Which “Side” Should Be Protected?

We learned that all enclosures should be protected. There are two questions that should be answered: (i) what is an enclosure? and (ii) should the enclosure be protected on the high-pressure side or the low-pressure side (or both)? These two questions are answered below. Enclosures can be divided into at least four categories:

- Pressure vessels
- Non-pressure vessels and tanks
- Pipes
- Casings of equipment.

The vessels and tanks are both types of “containers” and were discussed in Chapter 9.

An enclosure may face two types of pressure, both of which can be dangerous: pressure and vacuum. Some people refer to these as “internal pressure” and “external pressure” (Table 12.3).

Both overly high pressure and overly low pressure could be detrimental to an enclosure and the safety of the operators. For each of these four types, there are relief devices that prevent overpressure (which causes explosions) and relief devices that prevent under-pressure (which causes collapse).

PRDs protect an enclosure against pressure, whereas VRDs (vacuum relief devices) protect an enclosure against vacuum.

It would not be far off to say that the issue of overpressure (internal pressure) could jeopardize the safety of a sys-

tem to a greater degree than the issue of under-pressure (vacuum, or external pressure).

In contrast to pressure, which can go up to a few thousands kilopascals and more, vacuum can only go down to zero absolute pressure, or -101 kPag. Because of the “limited” nature of a vacuum, some designers may decide to get rid of a vacuum relief device via the system design; in other words, handling the vacuum issue with a passive solution rather than an active solution. A perfect way to do this is by designing an enclosure with a vacuum design of “full vacuum,” or FV. An FV enclosure ensures that the enclosure is protected against overly low pressure in the enclosure. However, designing an enclosure for FV could be very costly, especially for large enclosures. It is not common to design large tanks for FV as a design parameter; rather, FV may be a design parameter for smaller enclosures or vessels.

It is important to note that any value other than FV for a vacuum design parameter of an enclosure may still call for a vacuum safety device in addition to a pressure safety device.

The enclosures should be protected against both pressure (internal pressure) and vacuum (external pressure). However, if an enclosure is already designed for “full vacuum,” then it doesn’t need to be protected against vacuum. This is the case for plenty of pressure vessels.

12.6 Regulatory Issues Involved in PRVs

The need to install a pressure or vacuum relief valve may be based on technical and/or legal requirements.

Table 12.3 Two types of pressure in enclosures.

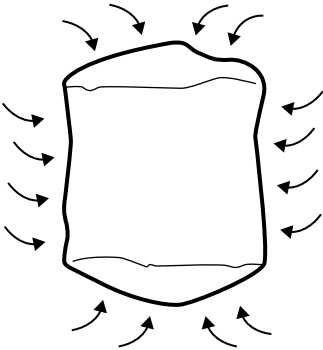
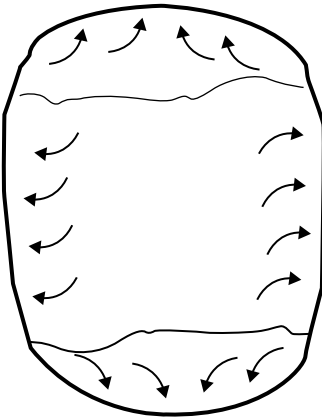
	Vacuum (external pressure)	Pressure (internal pressure)
Schematic		
Needs	Vacuum protection	Pressure protection

Table 12.4 Requirements for installing a relief device.

	Pressure vessel	Non-pressure vessel	Pipe	Equipment casing
Pressure	Regulated: legally required	Technically required	Technically required (in some areas this is regulated)	Technically required (not common)
Vacuum	Technically required (not required if FV)	Technically required	Technically required (not common)	Technically required (not common)

As a general rule, a regulatory body is involved when public health and safety is at play. The frequent explosion of boilers in the last century has brought high-pressure enclosures to the attention of regulatory bodies.

All in all, there are eight possible combinations of container type and pressure type, and each of them may need relief devices (see Table 12.4).

However, in terms of regulation, the most popular “legally” regulated case is that of pressure vessels with the potential for high pressure. This is because of their nature: the most catastrophic consequences (due to the size of the container) with the most catastrophic outcome (explosion). Other enclosures in other scenarios may also be overseen by a regulating body, depending on the region.

However, it is very important to mention that just because a case is not legally regulated, it doesn’t mean that it is not important and that there is no need for relief devices. A lack of rules simply means that, for the time being, the regulatory body doesn’t have enough resources to regulate all cases that may jeopardize safety, and the designer still needs to use his/her solid judgment to decide if there is a need for a safety device or not.

12.6.1 Codes Versus Standards

Codes and standards are both documents that are used by different parties involved in the PRD industry. However, there is a big difference between codes and standards. A code is a document generated by a regulatory body and it is mandatory that the people falling under that jurisdiction follow it. On the other hand, a standard is a document generally generated by a not-for-profit group of specialists like API (American Petroleum Institute).

The goal of a code is to protect the health and safety of personnel, as well as the environment. On the other hand, the goal of a standard is to minimize debates between different parties and reduce cost.

The people involved in a project may show a tendency to use standards or not. Following a standard is not mandatory, but many people try to follow them, usually because of a lack of a better and more reliable document.

Violation of a code in a jurisdiction will trigger legal action (after being revealed by an enforcement agent),

Table 12.5 Codes versus standards.

	Standard	Code
Conformance	Voluntary (but it is wise to do so)	Legal – mandatory
Goal	Reducing cost	Protecting the public
Originator	Not-for-profit group	Government or a government-authorized group
Enforcement agent	Doesn't exist	Government or a government-authorized group

**Table 12.6** Examples of codes and standards.

	Pressure vessel	Atmospheric tank	Pipe	Equipment casing
Standard	API 520, API 521	API2000	ASME B31.X	X
Code	ASME VIII-based BPV	x		x

while not following a standard doesn’t trigger legal action unless following the standard is mentioned in the contract (breach of the contract) or negligence occurs (tort action).

Table 12.5 outlines the differences between codes and standards.

Table 12.6 shows typical examples of codes and standards in the pressure relief device industry.

It is worth noting that regulatory bodies in different jurisdictions (different countries, states, provinces etc.) generally don’t generate a code from scratch. They generally start with a base document to save time and money. This document could be generated by a standards-developing organization and then be used as the “code template” by the regulatory body. Therefore, for each code we can talk about its “code basis”.

Table 12.7 shows typical examples of codes and standards in the pressure relief device industry.

Table 12.7 Codes in the pressure relief device industry.

	Code	Code basis	Enforcement agent (approval authority)
US (general)	NB-501, boiler and pressure vessel code	ASME B&PV code	NBBI (National Board of Boiler and Pressure Vessel Inspectors)
Canada (general)	Safety codes	CSA-B51, ASME B&PV code	A safety authority under the government: TSSA, ABSA, etc.

12.7 PRD Structure

The earliest type of pressure device was the one used for steam engines in the 1900s. It was basically a plug that exerted force using a hanging weight to close the pressure-releasing hole (orifice) and keep it close to the point that the pressure exceeds a specific value. The value of the pressure at which the device is intended to start to opening, or the “set pressure,” could be adjusted by sliding the weight along the lever (Figure 12.4).

Principally speaking, every pressure/vacuum relief device comprises three elements:

- 1) A pressure-sensing element
- 2) A logic
- 3) An opening element.

The pressure-sensing element senses and monitors the pressure of the enclosure and reports to the logic. The logic decides if/when the pressure exceeds a pre-set pressure, or set pressure, and sends orders to the opening element to open the device to enable the release of the excessively high pressure. In case of a vacuum, it can also “suck” from the outside of the enclosure to decrease (“break”) the vacuum in the enclosure (Figure 12.5).

Based on the fundamental concept of a PRD, all of these elements must be mechanical and their communication must be through mechanical links. Electrical, pneumatic, or hydraulic signals are not acceptable as communication routes in pressure/vacuum relief devices.

The above concept can be seen in the operation of a pressure relief valve (PRV) in Figure 12.6.

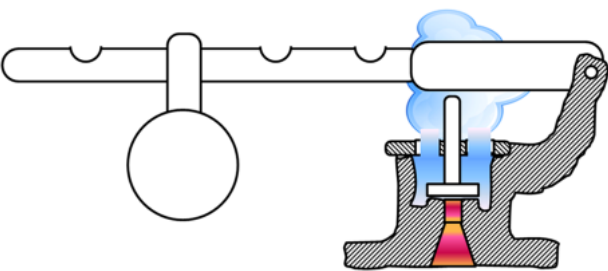


Figure 12.4 Early type of pressure relief valve.

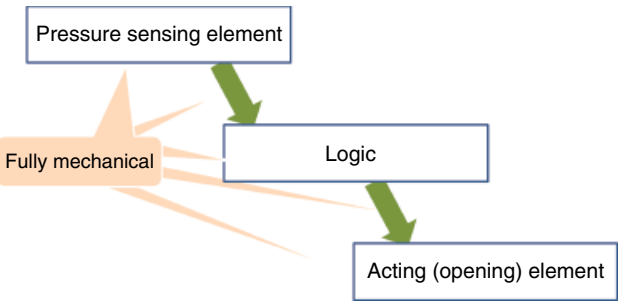


Figure 12.5 Fundamentals of relief device operation.

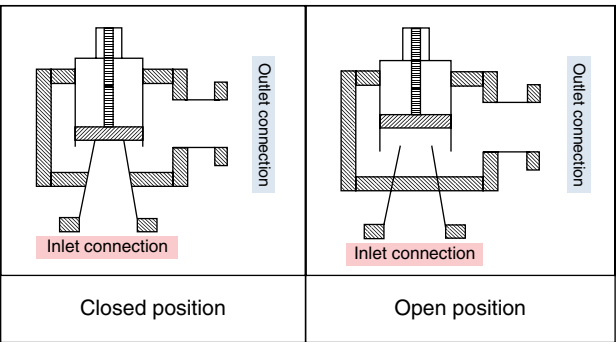


Figure 12.6 PRV schematic and operation.

12.8 Six Steps to Providing a Protective Layer

There are six elements that should be considered during the design stage to make sure an optimal protective layer is provided.

These elements are:

- 1) Locating the PRD
- 2) Positioning the PRD
- 3) Specifying the PRD
- 4) Selecting the right type of PRD
- 5) Selecting the right type of PRD arrangement
- 6) Checking the functionality of the PRD

We will discuss all of the above items except for items 3 and 4.

We briefly introduce “specifying PRD” because we need to understand the technical information of PRDs

on a P&ID. We however don't go to detail because this is a topic in the design stage of a project.

The last item won't be touched here at all again because it has nothing to do with P&ID development.

12.9 Locating PRDs

How do we know if a PRD is needed? Is there a legal obligation instated by your region's regulatory body? If so, then determining whether or not you need a PRD is easy: "We have to have it, we are going to install it." If not, then the question becomes whether there is a technical obligation to include a PRD in the design or not. Therefore, there are two questions you should ask yourself to decide whether a PRD is needed. These two questions are outlined in Figure 12.7.

From a purely theoretical point of view, a PRD is required if the following four criteria are met:

- 1) There is at least one valid overpressure scenario for a container with trapped fluid.
- 2) This overpressure scenario will increase the pressure of the container to higher than the maximum allowable working pressure (MAWP) during the life of the container.
- 3) This overpressure scenario does not void the integrity of the container prematurely, before PRD action.
- 4) The risk of explosion (without a PRD) is higher than what is tolerable.

However, from a practical standpoint, items 2, 3, and 4 are generally eliminated from the list of PRD requirement criteria. In the case of item 2, this is not always validated, irrespective of whether the overpressure criterion increases the pressure to higher than the MAWP or not; as long as there is one overpressure scenario (item 1) the requirement for a PRD is confirmed. However, in some cases where the ultimate pressure can easily and accurately be estimated, this criterion can be validated. One example of such a case is thermal expansion of trapped liquid.

For item 3, generally nobody checks if the integrity of the container is held intact before the opening of the PRD. This is because such an estimation would be difficult, if not impossible. If someone wants to challenge

the elimination of item 3 from the list, he may ask an exaggerated question such as: "do we need to install a PRD for a cardboard container?" Installing a PRD for a cardboard container may seem funny, but in reality, and for general materials of construction in the process industries, we generally install a PRD for enclosures irrespective of their material. So, for cases such as a fiberglass tank in a fire scenario, we still need to install a PRD.

For item 4, we don't go through the risk analysis to check whether the risk is higher than our tolerable level to see whether or not we need to install a PRD.

Basically, from a technical point of view, as long as there is a container with at least one overpressure scenario (meaning with trapped fluid – item 1), there is a need to install a PRD.

However, the regulatory body's view could be different from the purely theoretical viewpoint and the practical viewpoint. Some regulatory bodies have a practical view, but some of them have a stricter approach. Some regulatory bodies expect to "see" a PRD on every single container. Their view is basically: "if there is a container, there should be a PRD."

12.10 Positioning PRDs

If needed, where should a PRD be physically placed? Directly on the system to be protected, or far from it? At the top or bottom of the container? If it is not directly on the system to be protected, then what is the maximum allowable connecting pipe length? Which pipe fittings may be installed on this connecting pipe? Vertically or horizontally?

All of these questions about PRD installation are answered in this section, even though not all of these features are visible on P&IDs.

The short answer is that the PRD should be placed, whenever possible, directly on the system to be protected, vertically, upward, and at the top of the container (Figure 12.8).

To expand this short, simplistic answer it should be said that a PRD should be placed as close as possible to the protected enclosure, and as far away as possible from flow disturbance-causing elements such as elbows or static mixers. PRDs also should be protected from vibrations. Another important requirement is that a PRD should be easily accessible for inspection.

Now the question is, if – for whatever reason – we cannot put the PRD directly on the system to be protected, where it should be placed? The answer is that it can be installed upstream or downstream of the system or "on" the system through a connecting pipe. In all these case the connection between the to-be-protected system



1. Is it illegal not to include a PRD?	
2. Is it unwise not to include a PRD?	

Figure 12.7 Two questions regarding inclusion of a PRD for a system.

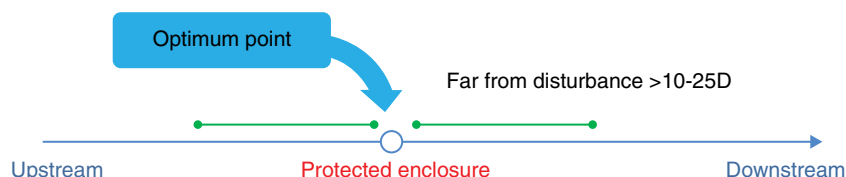


Figure 12.8 Positioning the PRD.

Table 12.8 Requirements of a PRD connecting pipe.

	Inlet	Outlet
Pipe	Pipe length minimized, pipe bore size maximized. Generally more critical in inlet of PRDs	Pipe length minimized, pipe bore size maximized Generally less critical in outlet of PRDs
Manual valve	Full port – CSO	Full port – CSO
Check valve	No needed	Rare, not a good practice
Control valve	Not permitted	Not permitted
Fittings	Minimize the number of fittings Generally more critical in inlet of PRDs	Minimize the number of fittings Generally less critical in outlet of PRDs
Reducer/enlarger	Reducer is allowable	Enlarger is allowable
Elbow	Long radius if the reaction pressure is large	Long radius if the reaction pressure is large
Tee	45°-wye instead of Tee if the reaction pressure is large.	45°-wye instead of Tee if the reaction pressure is large.
Pocket in pipe route?	“No Pocket”	“No Pocket”
Heat tracing and insulation	If needed, but could be very critical as PRD is placed on a “dead leg”	If needed

and the PRD is through a pipe. What are the requirements for this piece of pipe?

The requirement is that the connecting pipe should not reduce the release flow rate or increase the chance of blockage and consequently isolation of the PRD from the to-be-protected system.

As a general rule, the larger the bore, the shorter and straighter the pipe, and the fewer the fittings the better.

These requirements affect the length and bore size of the connecting pipe and also the fittings on it. A summary of such requirements are listed in Table 12.8.

An easy way to memorize the rules related to reducers and enlargers in pipe routes around the PRDs is memorizing this statement: “the connecting pipe cannot be smaller than the flange size.”

Figure 12.9 shows an example of a correct P&ID schematic of a pressure safety valve.

Placing the PRD at the top of the system (e.g. container) also eliminates the possibility of clogging due to the impure contents of the container.

The next issue to be resolved is whether the PRD should be installed vertically or horizontally. The answer depends on the type of PRD. Types of PRDs will be

discussed in Section 12.12.1 but in nutshell, there are mainly two types of pressure/vacuum relief devices: relief valves and rupture disks. PSVs should be installed vertically and upwardly. This limitation should be mentioned on the P&ID beside the P&ID symbol as a note. Sometimes the note of “install vertically and upwardly” could be disregarded where the skill of the pipe modelers is trusted. There is no such limitation for rupture disks (Figure 12.10).

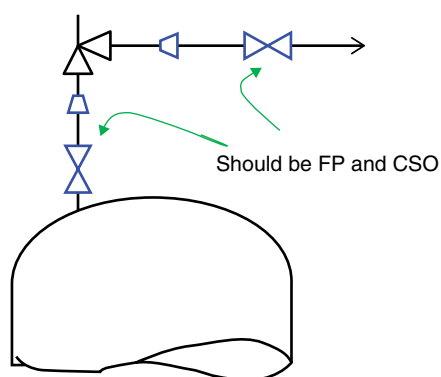


Figure 12.9 P&ID schematic of a PSV with connecting pipe.

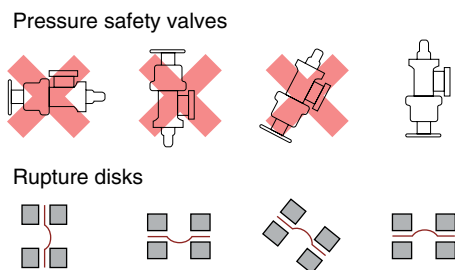


Figure 12.10 Vertical or horizontal installation of PRDs.

Table 12.9 PRD sizes.

	Safety valve	Rupture disk
Internal size	Orifice size in the form of standard designation letter or area of the orifice	Holder size
External size	Inlet and outlet opening sizes	Holder size

12.11 Specifying the PRD

Specifying a PRD goes to the design stage of a project. We only talk here about the result of such activity.

First of all a designer has decided on the pressure that the PRD should be opened to protect the enclosure. This pressure is named the “set pressure” for pressure safety valves and “burst pressure” for rupture disks.

Then the designer defines the different over-pressurizing scenarios, calculates the releasing rate for each of them, and finally picks the largest flow rate and names it the “governing scenario.” One famous over-pressurizing scenario is “fire.”

In the last step he specifies the PRDs by identifying the internal and external size of the PRD.

The internal size means the size of the hole inside of the PRD that the releasing flow goes through and the external size means the size (or sizes) of the inlet and outlet openings of the PRD.

The internal and external sizes for safety valves and rupture disks are stated in Table 12.9.

12.12 Selecting the Right Type of PRD

All pressure/vacuum relief devices can be classified into two main groups: reclosable devices and non-reclosable devices. The difference is that the former will open under excessively high pressure, and then will be reclosed when the pressure reverts back to normal pressure, whereas in the case of the latter, the device won’t reclose after its

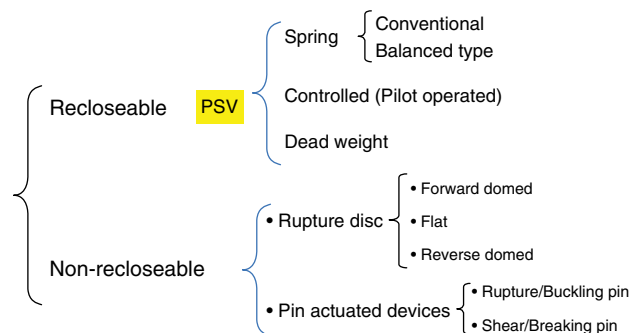


Figure 12.11 Different types of pressure/vacuum relief devices.

first and only opening. Basically, reclosable devices are non-disposable devices, and non-reclosable devices are disposable devices.

The reclosable devices are basically PRVs, and non-reclosables are breaking/rupturing element devices.

There are different types of pressure relief valves and breaking/rupturing element devices. A simple classification of all pressure/vacuum relief devices can be seen in Figure 12.11.

“Dead weight” PRDs were the first types of PRDs in industry and nowadays they are only used in non-pressure vessel applications.

Pin-actuated devices are not very common and are discussed here.

12.12.1 Pressure Relief Valve Type

The whole structure of a reclosable PRD is a force-exerting member with a pre-set force to keep a “hole” closed until the pressure of the enclosure goes beyond that pre-set pressure and push it back to open the “hole.” Therefore, a “pre-set force of a force-exerting member” is the heart of a reclosable PRD.

The “pre-set force of a force-exerting member” could simply be a disk with a pre-set weight; a dead weight-type, overpressure or a spring with a specific spring strength; spring-type, or a combination of internal pressure of the enclosure (tank or vessel) plus a (weaker) spring, or a pilot-type reclosable PRD.

Spring-loaded PRVs come in two different forms: conventional (with an uncovered spring) and the bellows type, which has a spring inside a bellows sleeve. The main difference between these two types of spring-loaded PRVs is that in the conventional type there is contact between the valve spring and the fluid inside enclosure, but in the bellows type the valve spring is isolated from the fluid by a bellows-shaped cover/sleeve.

This could be an advantage of the bellows-type PRV for cases where the fluid is aggressive and may impact

the functionality of the spring. The other feature of a bellows-type pressure relief valve is that the pressure downstream of the relief valve (or backpressure) doesn't impact the set pressure of the relief valve. Pilot-type relief valves also have this feature.

Schematics of the different types of PRDs are depicted in Figure 12.12.

12.12.2 Rupture Disks

There are two types of non-reclosable PRDs. In “rupture disks,” a “hole” is covered by a disk that will rupture at a specific pre-set pressure, and release the pressure. The second type of non-reclosable PRDs is very similar to a spring-loaded PRV, but the spring is replaced by a “buckling/breaking” pin. Between these two non-reclosable PRDs, rupture disks are more common.

Rupture disks are manufactured in main three forms: flat, forward dome, and backward dome. These three types of rupture disks could be in the form of a solid sheet, a hinged or scored sheet, with a cutting edge, and in composite. The available types of rupture disks are shown in Table 12.10.

12.12.3 Decision General Rules

Deciding on PSD types are based on quantitative and qualitative parameters. As many criteria go back to the design stage of project, they are not discussed here.

12.13 PRD Identifiers

As it was stated in Chapter 4, the identifiers of PRDs – as an item of instrumentation – on P&IDs are PRD symbols, PRD tags, and PRD technical information.

12.13.1 PRD Symbols and Tags

Table 12.11 shows P/V RD symbols and tags.

As can be seen in the table, when there is a need to have both a pressure relief valve and a vacuum relief valve, these two devices can be merged together to save money on nozzles and other operating costs. The PVSV was invented for this purpose. PVSVs (pressure/vacuum safety valves, or as some companies call them, PVRVs [pressure/vacuum relief valves]) are devices that protect

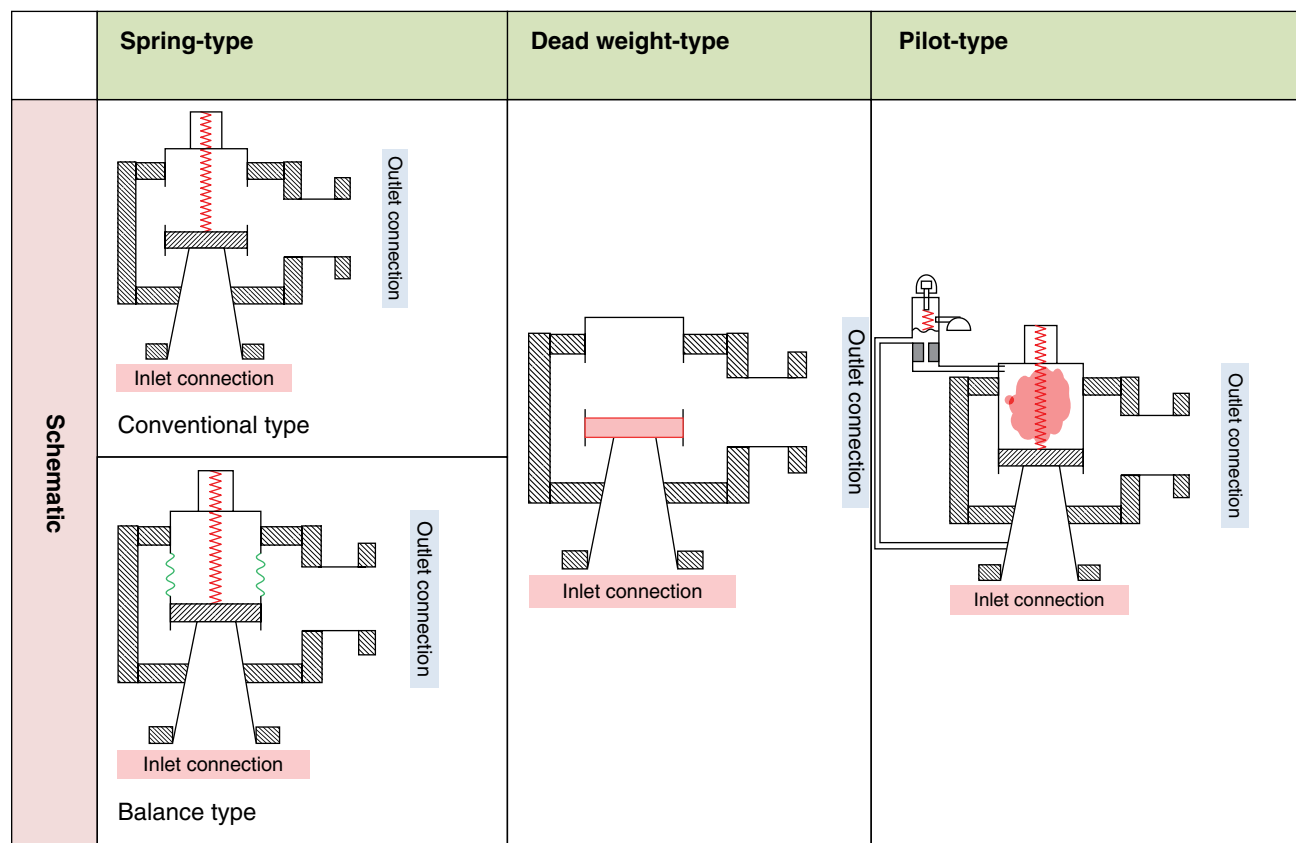


Figure 12.12 Different types of relief valves.

Table 12.10 Different types of rupture disks.

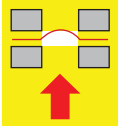
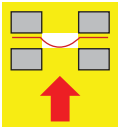
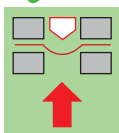





	Solid (standard)	Scored, hinged	With cutting edge	Composite
Flat	Old usage, still available in low pressure HVAC applications	✗	✗	The only common flat rupture disk these days
Forward dome	✓	✓	✗	Available
 (Pre-bulged, forward acting)				
Reverse dome	✗	✓	✓	✗
 (Reverse acting)				

Table 12.11 P&ID symbols for different relieving devices.

	Safety valve	Rupture disk
Pressure	 PSV	 PSE
Vacuum	 VSV	 VSE
Combination	 PVSF	Not available

a container (generally tanks but not vessels) against both pressure and vacuum.

12.13.2 PRD Technical Information

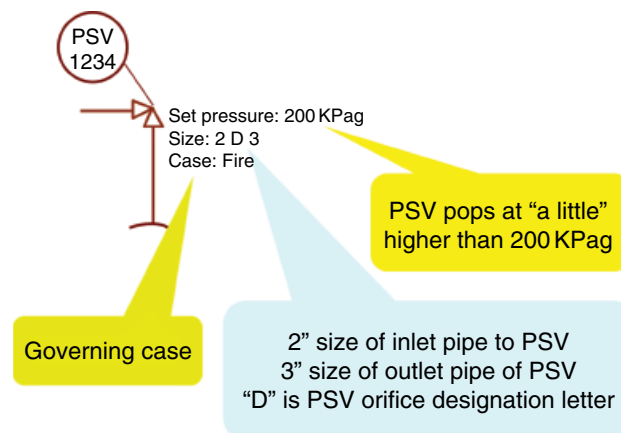
The technical information about the pressure relief devices are placed near their symbols on P&IDs. However some companies prefer to insert such technical

information as a “call-out” on the top or bottom of the P&ID sheet.

The first important technical information for PRDs is the “set pressure”, for PSVs, or “burst pressure,” for rupture disks.

The second important piece of technical information is the PRD size.

Recently more companies prefer to state the “governing case” as the other piece of information near the symbol of PRDs on their P&IDs.

**Figure 12.13** PSVs on P&IDs.

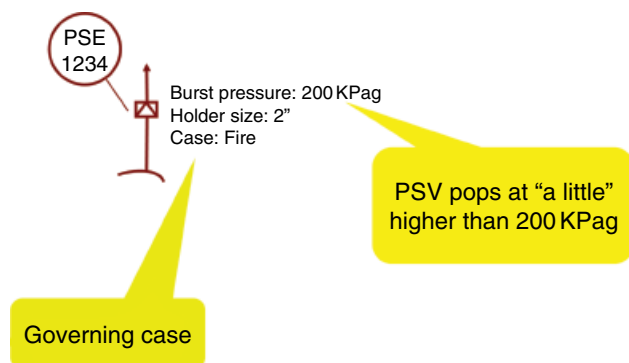


Figure 12.14 Rupture disks on P&IDs.

The sample technical information of a PSV is shown in Figure 12.13 and for a rupture disk in Figure 12.14.

12.14 Selecting the Right Type of PRD Arrangement

The PSDs could be installed in different arrangements. The simplest arrangement is a single PSD. Multiple PSDs could be placed in some cases. They could be in series or in parallel.

When they are in series they could be the same type or different types. In the series arrangement it could be a PSV and a rupture disk or two rupture disks in series.

When they are in parallel arrangement they could be all functional (such as a $2 \times 50\%$ or $3 \times 33\%$ arrangement) or a functional-spare arrangement (such as the $2 \times 100\%$ spare philosophy).

The parallel arrangement can also be classified based on the similarity or dissimilarity in types. There could be multiple PSVs in parallel, or multiple rupture disks in parallel, or few PSVs and few rupture disks in a parallel arrangement.

The different arrangements of PRDs can be seen in Figure 12.15.

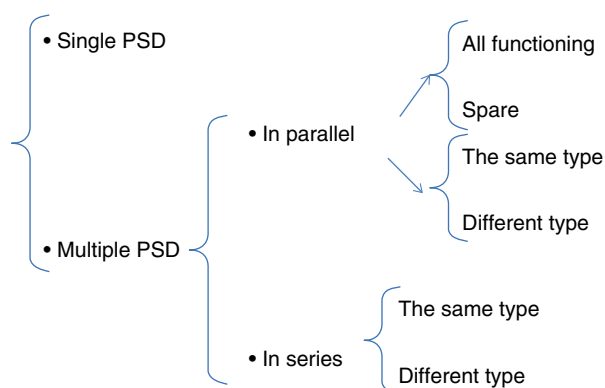


Figure 12.15 Different PRD arrangements.



Figure 12.16 A single simple PRD.

The single PSV is the default choice to protect process enclosures (Figure 12.16). The other option is using a single rupture disk.

However if there is a need to inspect or maintain the PRD while the rest of the plant is operating, the PRD needs an isolation system (isolation valve and drain/vent valves) and some other provisions.

The provisions are the systems to allow pulling the PRD out of operation and to allow an operator to function as a “PRD” for the time the PRD is out of operation. There should be a connection to connect a portable pressure gauge (or an already installed pressure gauge in place). This system should be completed with a bypass pipe and a manual throttling valve on it. During the time the PRD is isolated from the to-be-protected system, an operator needs to watch the pressure gauge and also be ready to open the manual throttling valve to release the pressure if the pressure goes higher than allowable.

Such a “manual safety system” is shown in Figure 12.17.

However not in all cases such “simple” system can be used. The other systems to support inline care for PRD’s will be discussed later here.

There are some cases that a single PSD doesn’t satisfy the requirement of safety. In such cases PSDs in parallel or in series are used.

PSDs can be placed in parallel for different reasons.

A non-exhaustive list of cases in which we may need to use parallel PRD’s is:

- When the required PRD is bigger than the maximum available PRD in the market and several smaller PRDs in parallel and all functioning are required.
- When the required PRD is big and for economical reasons it is better to use several smaller PRDs in parallel and all functioning.
- When there is a difference of more than 20–30% of release flow in different credible scenarios, two (or more)

Figure 12.17 A PRD with provisions for inline care for the PRD.

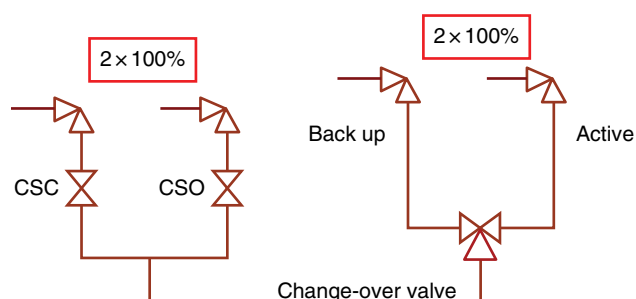
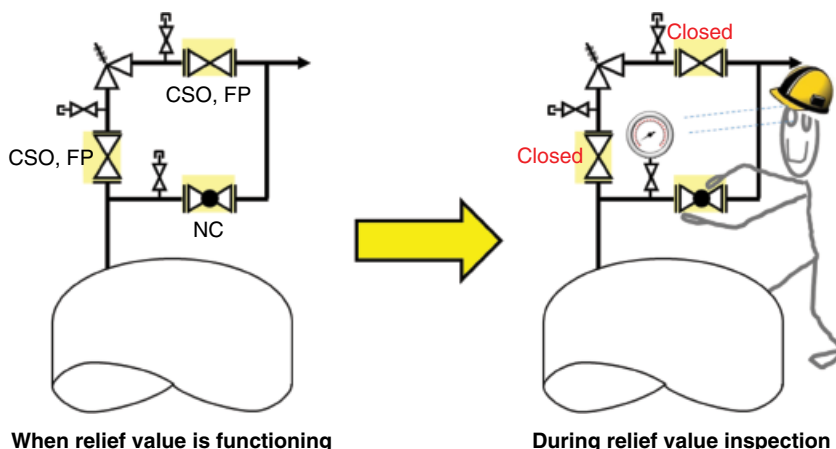


Figure 12.18 2 x 100% spare PRDs.

PRDs with different sizes in parallel and all functioning should be used.

- When one PSV doesn't provide the required "probability of failure on demand" for the system two (or more) PRDs with different sizes in parallel and all functioning should be used.
- When more than one scenario is valid and each of them calls for a different type of PSD two (or more) PRDs with different types in parallel and all functioning should be used.
- When one PSV cannot satisfy the required reliability of the system. In this case, again, we are going to have sparing philosophies like 2 x 50% or 3 x 33% for the parallel PRDs. The spare PSDs are used where there is a need to inspect, recalibrate, and repair a PSD during plant operation and the "manual safety system" is not acceptable.

When there are spare PRDs each of them should have isolation valves. The critical items are the isolations valves upstream of each PRD. The functioning PRD should have car seal open CSO attributes and the spare PRD car seal close (CSC). We have to make sure that when one upstream isolation valve is open the other one is closed. Sometimes this is done by placing one "change-over valve" to make sure no mistakes happen (Figure 12.18).

PSDs can be placed in series mainly for one reason: to isolate the main PSD from the aggressive or dirty process fluid. In this arrangement a rupture disk is almost always placed upstream of (below of) the main system protector, which could be a PSV or a rupture disk (Figure 12.19(a)).

If there is rupture disk leakage the main purpose of the rupture disk is missed; the connecting pipe of rupture disk to the PSV should be "checked."

As there is always a chance of rupture disk leakage that the operators are unaware of the "space" should be monitored.

The most common solution is installing a pressure gauge on the space. A more advanced solution is installing a pressure switch or pressure sensor to warn of the increase in pressure (because of the rupture disk leakage) to the control room (Figure 12.19(b)).

The other solution is to connect the rupture disk to a burst sensor. The burst sensor sends a signal to the control room in the case of rupture in the rupture disk (Figure 12.19(c)).

In Figure 12.19(d) a pressure alarm warns the operators if high pressure is created in the pipe between the rupture disk and the safety valve.

For more severe leakages a pipe can be connected to the rupture disk-PSV pipe. On this piece of pipe there should be some systems. One example is placing an excess flow valve on it (Figure 12.20).

Each of above solutions has advantages and disadvantages and one may decide to use a combination of them (Figure 12.21).

In less severe cases where the fluid is only mildly precipitating, a "flush ring" can work perfectly without the need to go to the more complicated case of using the combined rupture disk and PSV (Figure 12.22).

If there is a chance of contacting aggressive fluid on the downstream of the PSV, there could be another rupture disk installed on the PSV outlet (Figure 12.23).

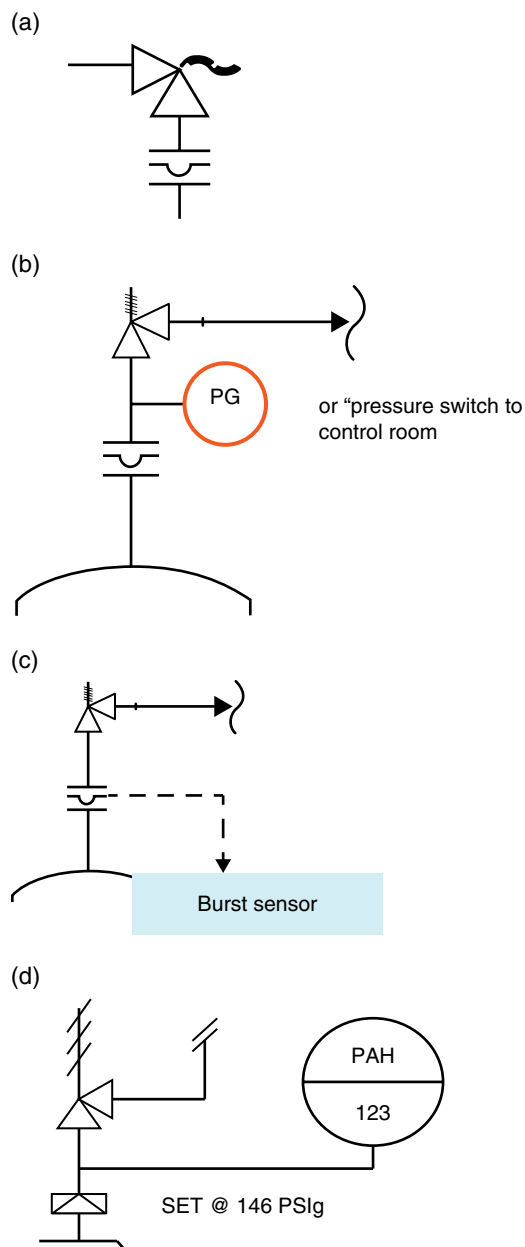


Figure 12.19 Combination of safety valves and rupture disks.

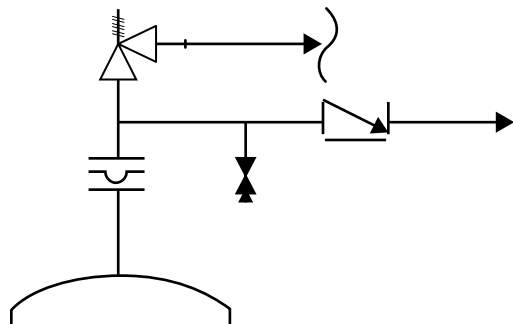


Figure 12.20 Combination of safety valves and rupture disks.

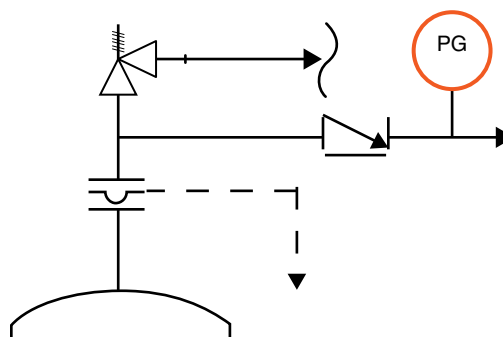


Figure 12.21 A combined solution to deal with leakage of the rupture disk upstream of the PSV.

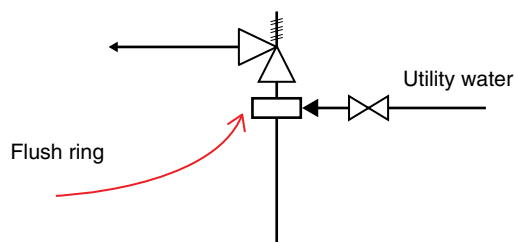


Figure 12.22 A PSV with a flush ring.

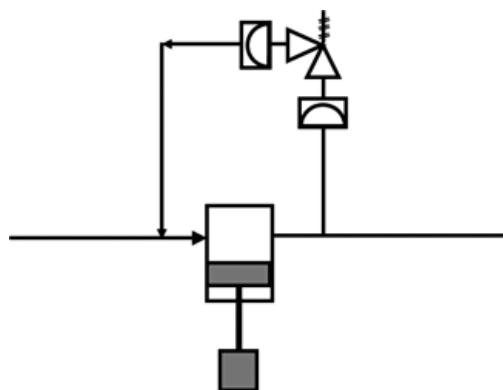


Figure 12.23 Rupture disk upstream and downstream of a PSV on the outlet of a PD pump.

12.15 Deciding on an Emergency Release Collecting Network

One important issue regarding P/VRDs is the final destination of the released fluid. An emergency release collecting network could be used to collect the instantaneous releases from PSDs and direct them toward the disposal system.

Figure 12.24 shows a simple schematic of an emergency release collection network.

The emergency network has several requirements, which are shown in Figure 12.25.

As is visible from Figure 12.25, the main header should be sloped toward the disposal system. If there is needed

Figure 12.24 An emergency release collection network.

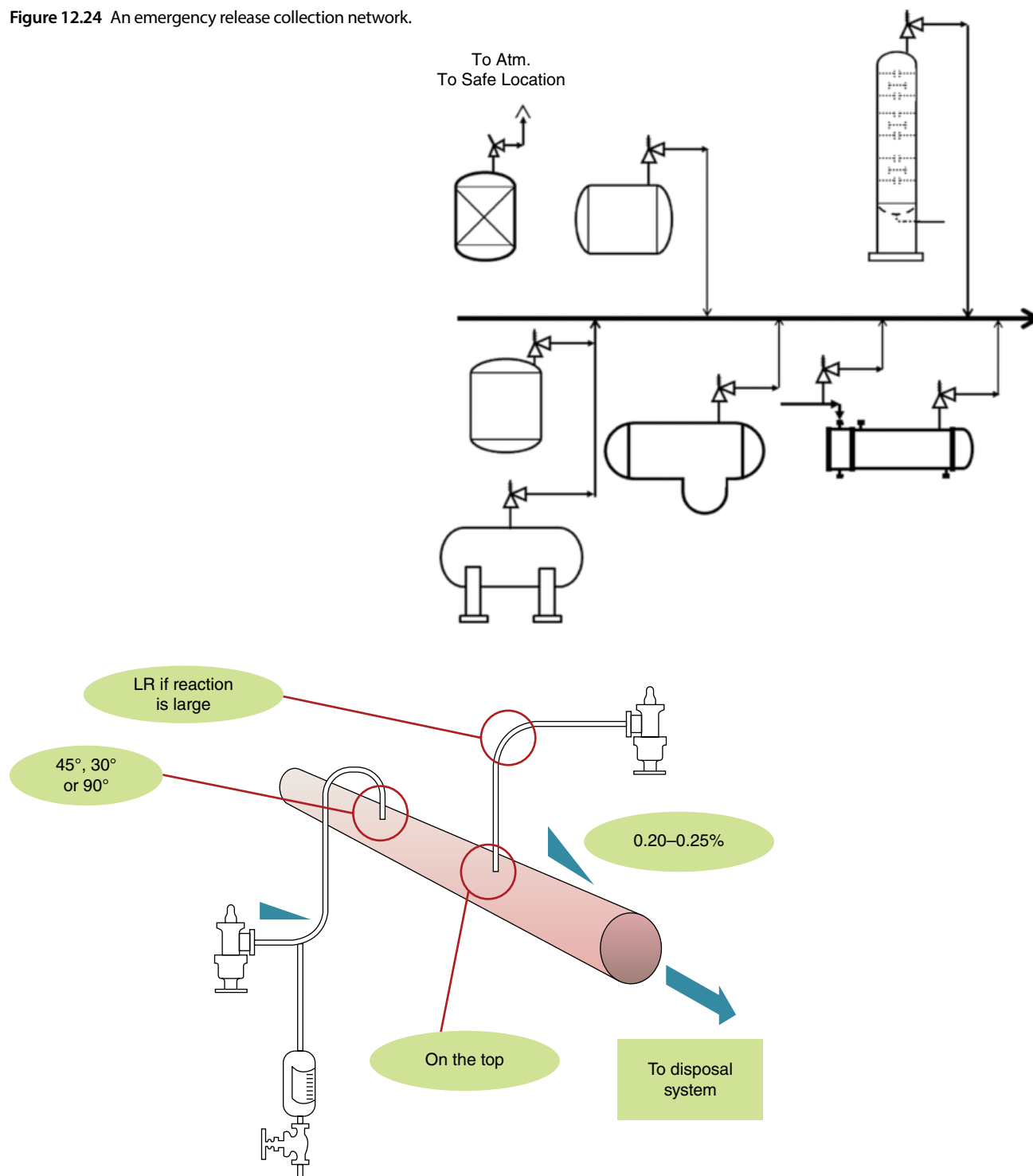


Figure 12.25 Features of an emergency release collection network.

to put valve(s) anywhere on the network, such as a “root valve”; it should be “CSO”.

The outlet of the PRDs should tie-in from the top of the header to make sure no liquid comes back and fill the outlet of the PRDs.

As in the outlet size of PRDs the “no pocket” rule is applicable, which means the PRDs should either be above the header or equipped with a system to consciously or intermittently drain the collected liquids from the outlet pipe of PRDs. This system should prevent escape of gas and vapors.

This system could be as simple as a one $\frac{3}{4}$ " drain hole on the elbow (Figure 12.26) to a manual draining pot (Figure 12.25) and to a "drip pan elbow" (Figure 12.27).

In the cases that releasing fluid may generate a huge reaction on the piping system, it may be decided to use a Y-fitting instead of a T-fitting, and long radius elbows instead of standard elbows. This could be the case when

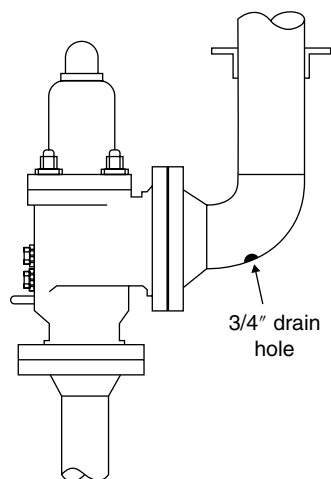


Figure 12.26 Preventing liquid accumulation by using a drain hole.

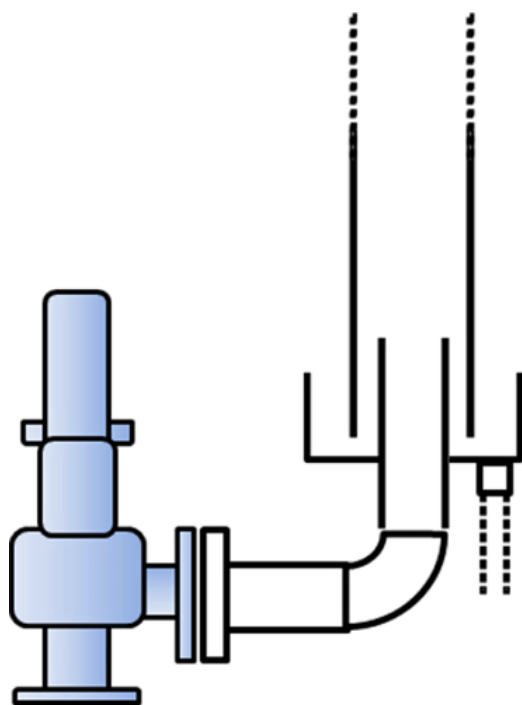


Figure 12.27 Preventing liquid accumulation by using a drip pan elbow.

the releasing flow rate is huge and/or the set pressure of the PRD is very high. Such requirements could be captured as notes on the P&ID.

12.16 Deciding on a Disposal System

The decision on the disposal system of choice depends on different parameters. The parameters are fluid phase and type, volume of released fluid, the governing overpressure scenario, plant constraints, and environmental regulations.

The first thing is whether the released fluid phase is liquid or gas/vapor.

The other parameter how "innocent" is the released fluid? If the release gas is an air pollutant or not, if the release liquid is flammable or not, etc.

The third parameter to consider is the volume of released fluid. It is important to notice the released fluid volume is different from the released flow rate. The flow rate can be huge but if it happens in a short period of time, the volume will be small.

A decision on the selection of disposal system is sometimes made during the PFD development stage and sometimes during the P&ID stage of the project. However, here we cover them briefly (Table 12.12).

12.16.1 Liquid Disposal

For liquid release there are different ultimate destinations available; however, "system relieving" is always the first choice. System relieving means sending the released liquid somewhere inside of the process units. Even though it looks like system relieving "should" be the first choice for gas relieving too, this it is not the case. Even though we love system relieving because the released fluid goes back directly to the process with no wastage, for gas release it is generally hard to find a part of the plant that is large enough to be able to contain the expanding high pressure gas stream.

System relieving is a common solution for PSVs on PD pumps (Figure 12.28(a)).

The other example of liquid system relieving is in PSVs to protect pipelines in thermal expansion scenarios (Figure 12.29).

The next option is relieving the liquid to the sump. The sump could be of open drain type of closed drain type (Figure 12.28(b) and (d)).

Another option is relieving the liquid to a small bucket (Figure 12.28(c)).

In older days it was common to spill the relieved liquid on the ground, but these days because of environmental regulations it is hard to do that.

Table 12.12 Ultimate destination.

Liquid	Gas/vapor
<ul style="list-style-type: none"> • System relieving • Open drain • Closed drain • Small vessel on the ground • Ground 	<ul style="list-style-type: none"> • Atmosphere (if innocent) • Flare (if combustible) • Emergency scrubber or quench pool (if contains absorbable and non-innocents) • VRU (if recoverable) • System relieving

12.16.2 Gas/Vapor Disposal

The first choice for relieving gas/vapor is disposing of it in the atmosphere because it is the least expensive choice!

These days this choice is allowable only if the relieving gas/vapor is innocent.

Disposing to atmosphere is generally regulated in different jurisdictions. The regulatory bodies allow release to atmosphere if the release doesn't jeopardize the health

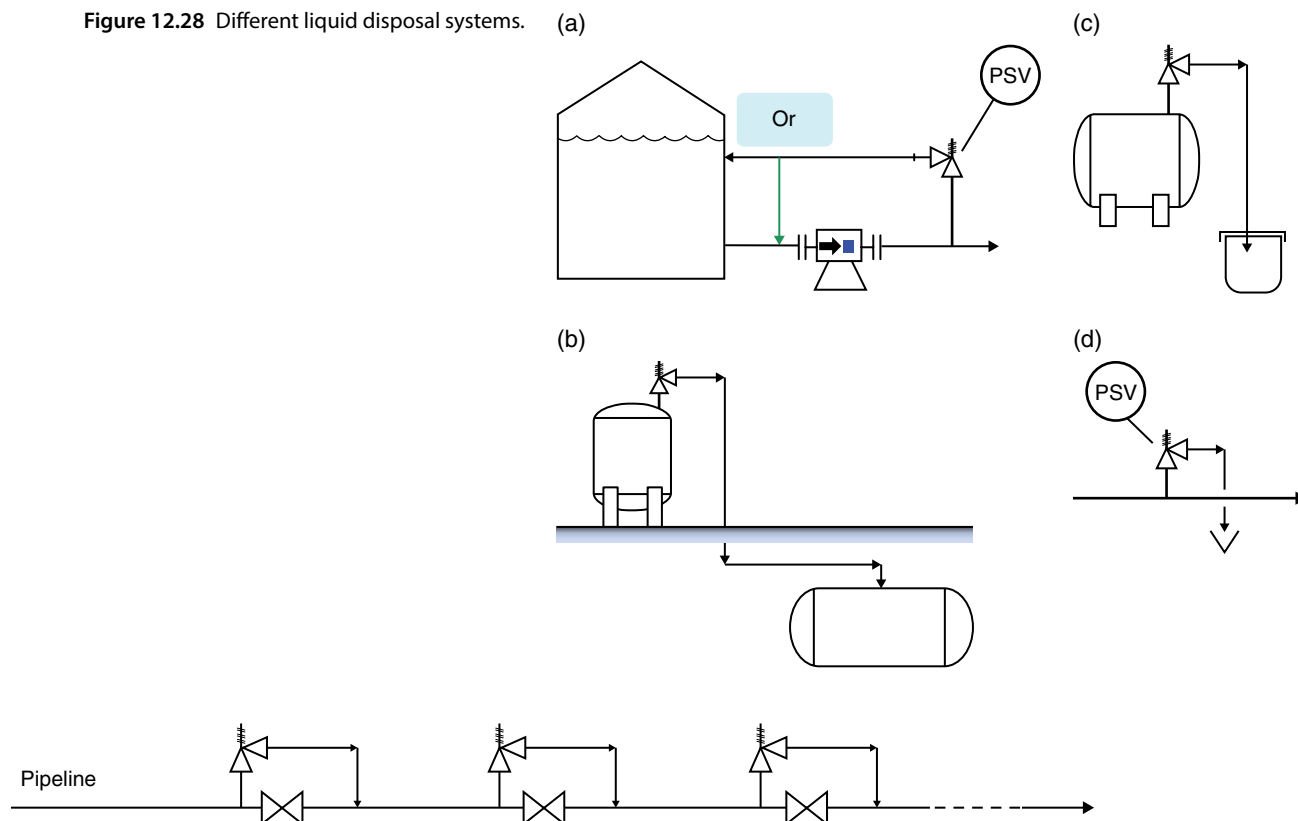
and safety of people and doesn't generate pollution in the environment.

The famous examples of releasing to atmosphere are releasing air or steam to the environment.

Wherever it is allowed to release to atmosphere a note is placed on the P&ID as "to atmosphere." The other note, which could be confusing, is "to safe location." Although a P&ID developer may not care about the meaning of this note, he needs to know this as there are some cases that finding a "safe location" is not easy!

Each company may have their own interpretation of "safe location for PRD release" and generally their interpretation is outlined somewhere in their guidelines. However, where there is no guideline the "safe location" can be assumed as: vertical and upward release so that the release point is higher than a minimum 3 m from the ground (or platform) and a minimum 2 m from the top of all equipment with a radius of 7.5 m around the release point (Figure 12.30).

The outlet of the PRD to atmosphere may need to be equipped with a "bird screen" to prevent bugs and larger animals from entering. Bird screens are tagged as an SP item on P&IDs.

Figure 12.28 Different liquid disposal systems. (a)**Figure 12.29** System relieving for thermal expansion PSVs on a pipeline.

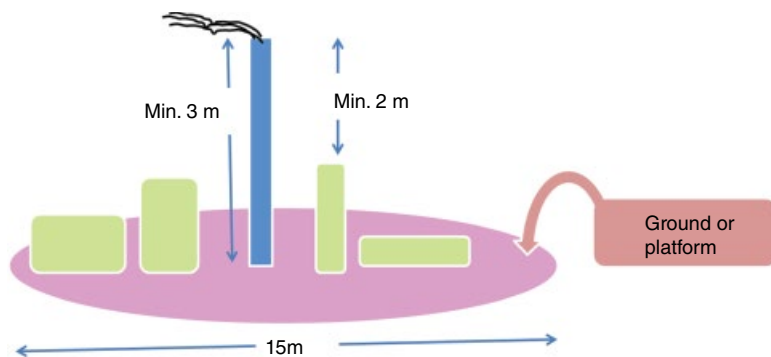


Figure 12.30 General meaning of “safe location” for releasing to atmosphere.

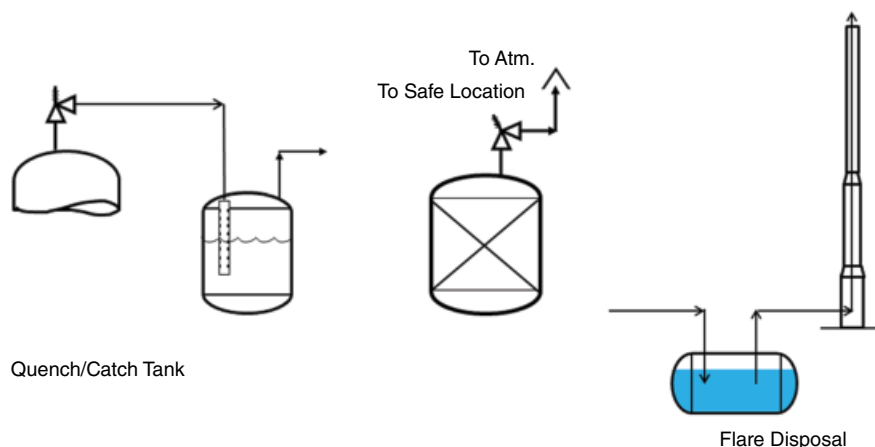


Figure 12.31 Different gas/vapor disposal systems.

If the relieving gas/vapor is not innocent, other, more expensive, options should be considered.

The options depend on the nature of the released gas/vapor; if it is flammable it can be burnt in a flare; if it is absorbable in water the stream can be sent to a catch vessel.

Figure 12.31 shows the different available options for gas/vapor relieving.

The last choice for gas/vapor relieving is “system relieving.” As was mentioned before, system relieving is not common for gases and vapors. However there are some cases that no other option is available and system relieving is the only technically doable option, for example in the oil extraction industry. Well pads are not necessarily close to the central plant, which has a flare system. There are, however, some units on well pads that have PRDs. Sometimes a large “pop tank” is located to release gases from the PRDs on the well pad. Pop tanks can be used for liquid relieving too.

12.16.3 Two-Phase Flow Handling

Here two-phase flow refers to gas–liquid two-phase flow.

We generally don’t provide a “two-phase flow disposal system.” What we try to do is to separate the two-phase flow to its components, gas and liquid, and then deal with each of them separately.

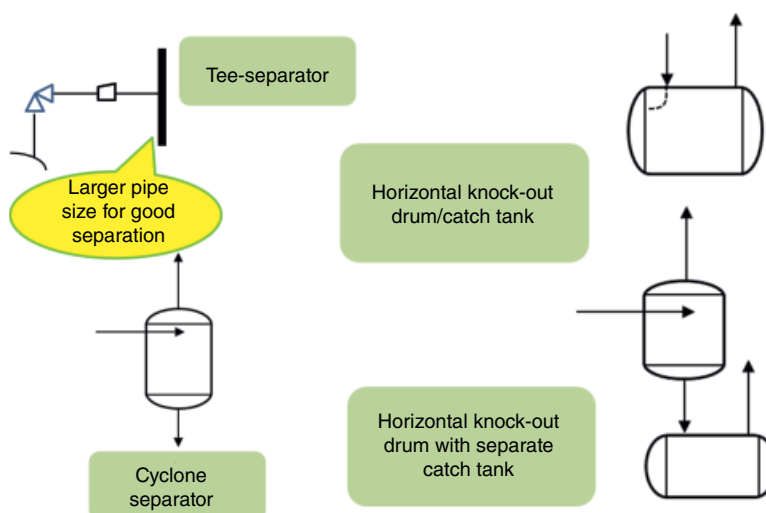
The reason for the separation of two-phase flow is two-fold. On the one hand the design and fabrication of a two-phase flow collection network is more complicated and expensive. On the other hand it is not easy to find a disposal system suitable for both liquid and gas at the same time.

However, if the liquid fraction or gas fraction of two-phase flow is very small, the two-phase flow can be considered to be a single-phase flow.

The two phase separators could as simple as Tee-separators, to the more complicated options of cyclone separators and knock-out drums (blow-down drums) (Figure 12.32).

Sometimes even a combination of them is used.

Figure 12.32 Emergency release two-phase separators.



12.17 Protecting Atmospheric Containers

Atmospheric containers are not only atmospheric tanks, they could be vessels that are designed with a design pressure of less than 15 psig. One example is low pressure evaporators, which are used to produce potable water from seawater. These evaporators are generally vertical vessels but their design pressures are very close to atmospheric pressures.

For atmospheric tanks all the overpressurizing scenarios could be divided in two groups of normal overpressurizing scenarios and emergency overpressurizing scenarios. A normal overpressurizing scenario is caused by normal breathing of tanks when the system designed to deal with it has failed. It is a good idea to deal with normal breathing separated from other overpressurizing scenarios. The reason is that it happens in almost all types of tanks and the other reason is that the relieving or sucking of vapors for breathing is huge.

On the other hand it was stated before that there are two pressures that enclosures should be protected from: internal pressure and external pressure (vacuum).

Therefore there are four components that we need to deal with regarding tanks (Table 12.13).

Although there are four components but we don't need to put four separate safety valves to address them.

The atmospheric containers should be protected against overpressure. As they are generally not designed "full vacuum" they also need to be protected against underpressurize or vacuum. One valve is used to cover vacuum conditions in normal and emergency cases.

One safety valve is used to deal with emergency overpressurizing scenarios and one (or more) safety valves to deal with normal overpressurizing scenarios.

The emergency valve is called a "pressure safety valve" (tagged as PSV). This valve is to protect the tank against emergency cases like fire.

The vacuum safety valve – for both normal and emergency underpressurizing scenarios – is tagged as VSV and the pressure safety valve for normal overpressurizing scenarios is tagged as PSV.

To save the number of nozzles on tanks, these two valves could be combined together into one valve called a "breather valve" or a "conservation valve" (tagged as PVRV or PVS), which is placed on one nozzle. The symbol of a breather (conservation) valve is shown in Figure 12.33.

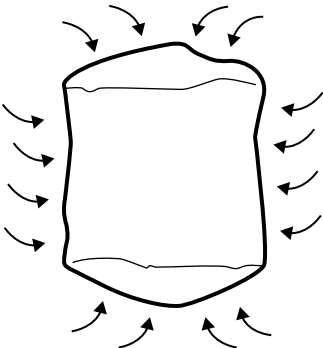
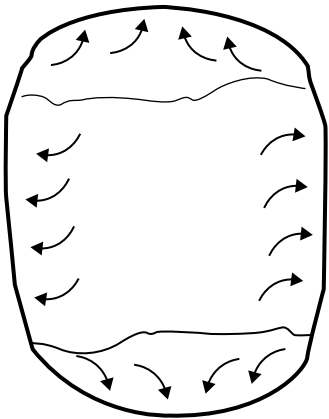
The nozzle sizes of these valves are typically 16", 18", 20", or 24".

Table 12.14 summarizes the safety valves required for vessels and tanks.

Table 12.13 Four components of pressure issues in tanks.

Pressure			Vacuum
Normal	Reason	Because of normal breathing; inhalation phase	Because of normal breathing; exhalation phase
	Extent	The pressure can go as high as possible	Vacuum has a limited extent, down to -101 KPag
Emergency	Reason	Because of emergency cases like fire	Because of emergency cases
	Extent	The pressure can go as high as possible	Vacuum has a limited extent, down to -101 KPag

Table 12.14 Two types of pressure in enclosures.

	Vacuum (external pressure)	Pressure (internal pressure)
Schematic		
Needs	Vacuum protection	Pressure protection
Pressure vessels	PSV, mandatory	Vacuum breaker • No needed if the container is designed "FV"(full vacuum)
Atmospheric tanks	Emergency Emergency vents or PSVs	Normal PRV Could be merged as PVRV or PVSV

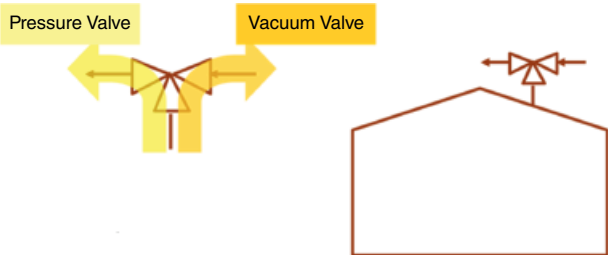


Figure 12.33 PVRV symbol.

The other, less common, way to protect non-pressure vessels are “seal locks.” Seal locks are a U-shaped pipes filled with water (or a freeze resistance liquid) in it (Figure 12.34). The height of the liquid column works in a similar way to a “dead weight” in dead weight PRDs. The problem of such PRDs is the necessity of frequent inspection of the seal lock to make sure the liquid is in working conditions.

12.18 Merging PRDs

Merging PRDs can be discussed in two contexts: merging two (or more) PRDs from different enclosures and

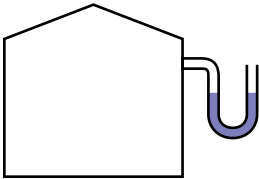


Figure 12.34 Seal lock.

putting a single PRD instead and merging two nozzles on a container and using a dual purpose device instead.

Let’s start with the first topic as it is a heated topic because it is of a technical and legal nature. In a nutshell, the PRDs of two separate enclosures can be merged together and only one PRD is used if there is no “obstacle” between the two enclosures.

In Figure 12.35 a “story” is presented to explain this concept. Let’s start the story from a purely technical viewpoint (not legal).

In Figure 12.35(a), a single enclosure exists and it has its own PRD.

In Figure 12.35(b), the above single enclosure is turned into an enclosure with two compartments. There is narrower enclosure between the two compartments.

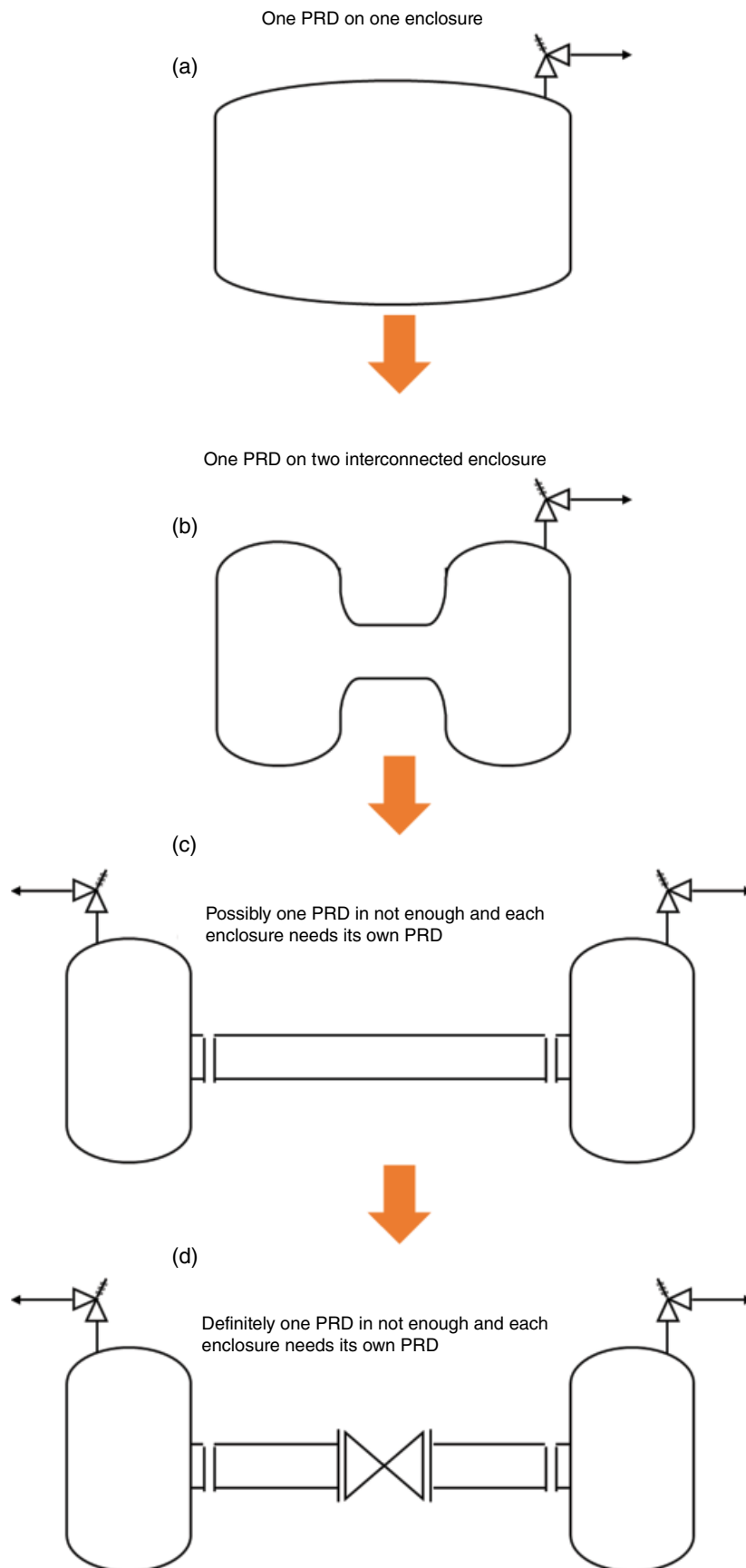


Figure 12.35 Evolution of one enclosure to two enclosures and the concept of a PRD.

A single PRD may still be adequate for this condition. The second PRD on the second compartment may be needed if there is a chance of plugging the “throat” enclosure and create two enclosures.

In Figure 12.35(c), the above single enclosure is turned into two interconnected enclosures. A single PRD is most likely not adequate and each enclosure may need their own PRD. A second PRD on the second enclosure may be needed if there is a chance of plugging the interconnecting pipe, or blocking the pipe because of a carelessly left behind blind after a maintenance activity, and create two enclosures.

In Figure 12.35(d), the above two interconnected enclosures have a valve on the interconnecting pipe. A single PRD is definitely not adequate and each enclosure should have its own PRD.

However the situation could be stricter from a legal viewpoint. The legal regulatory body may force you to install two PRD’s, even for cases Figure 12.35(b) and (c).

The second topic is using a dual purpose device instead. There are efforts to merge safety valves for the normal overpressurizing scenario with other nozzles on tanks. The reason is they are large nozzles and a tank may need several of them. As the size of such safety valves are in the ball park value of thief hatches and manways the main efforts resulted in inventing a “thief hatch plus safety valve” and a “manway plus safety valve.” Figure 12.36 shows one of them.



Figure 12.36 Combined PSV and thief hatch; “gauged thief hatch.”

12.19 Wrapping-Up: Addressing the Requirements of PRDs During their Lifespan

In this section we check our design to make sure we cover all the needs of PRD’s during each phase of plant life. As was discussed before these phases are normal operation, non-normal operation, inspection/maintenance, and operability in the absence of one item.

- 1) Normal operation of pipe: a PRD actually functions during the emergency situation in a plant. The required considerations are already covered.
- 2) Non-routine operation (reduced capacity operation, start-up operation, upset operation, planned shut-down, emergency shut down): PRD invented to deal with such situations.
- 3) Inspection and maintenance: generally speaking PRDs need much attention during the plant operation. General consideration regarding inspection and maintenance of all items is covered in Chapter 8.
- 4) Running the plant in the absence of pipe: This is never acceptable!

Part III

Instrumentation and Control System

There are four “tools” that are used to control and operate a process plant smoothly. They are the automatic control system, safety interlock system, alarm, and mechanical relief.

These systems help operators to run a plant. Out of these four tools, mechanical relief was already discussed in Chapter 12. There are three tools left to be discussed.

Before discussing the control system some preliminary information such as terminology and symbology should be discussed, and these are covered in Chapter 13.

Control systems are discussed in Chapters of 14 and 15.

Chapter 14 talks about important control loop architectures. Chapter 15 is about process plant control and has two parts: plant-wide control and equipment-wise control.

Chapter 16 is devoted to safety interlock systems and alarms. A small portion of control systems is also covered in Chapter 16. This portion is discrete control and is introduced in Chapter 16 because of its similarity with safety interlock actions.

Finally, in Chapter 16, control and interlock of electric motors are discussed as an important example.

To explain the control concepts not necessarily the terminology of academia or even industry is used. The goal is conveying the concepts in the simplest way.

It is important to be stated that the schematics in this part are provided not as detailed as the rest of this book. The schematics here are mainly to convey the concept of control system and then some mechanical details are ignored.

13

Fundamentals of Instrumentation and Control

13.1 What Is Process Control?

To explain a control system, we can take an example from everyday life.

A father asks his daughter to avoid the edge of a cliff while they are hiking. This first step is a “regulatory” measure. A while later, the father sees his daughter playing at the edge of the cliff; this raises an alarm and the father tells her more harshly to be careful and step back. If the daughter continues to play by the edge, then the father may decide to take control of the situation; he may take her by the hand so that they can both leave the park. These three steps relate closely to a typical control system.

Similar to the above analogy, the first step in a process control system is “regulation.” Regulatory control is exercised to ensure that the process runs smoothly and according to specifications. The more technical term for this is a “basic process control system” (BPCS).

If the process runs out of control or off-spec, then an alarm is raised. This is the second step in the control system, which alerts the operator of the need to implement stronger action to correct the situation.

If the out-of-control condition persists, then the control system moves to the third step, which may involve drastic action. In plant process control, this is called a safety instrumented system (SIS).

So when we talk about a process control system, we generally refer to three elements: BPCS + alarm + SIS. Collectively, these are called an integrated control and safety system (ICSS). We need to be specific when referring to any particular system because the general term “process control system” may mean a BPCS to some people and an ICSS to others.

Manipulating a plant is not limited to ICSS. ICSS is basically “automatic control,” but we still need operators’ presence in plants. What operators do is basically “manual control.”

They should be in the field and in the control room to monitor parameters to take actions when they see an emergency case.

Over the life of a plant, there will be what we can think of as “sunny days” and “rainy days.” On sunny days, the process runs smoothly, there is no threat to the system, and everyone is happy. On sunny days a plant is run by a BPCS system. Rainy days are when there is one or more upsets in the plant. On rainy days, a plant is run by SIS actions. SIS actions cannot be manual, but operator intervention is provided for manual interference in automatic SIS actions (Table 13.1).

13.2 Components of Process Control Against Violating Parameters

As was mentioned in Chapter 5 there are four steering/protecting components for each specific process parameter in equipment, units, and plant.

These steering/protecting components are:

- BPCS. The main function of the Basic Process Control System is to ensure that the plant runs smoothly and within specifications. This is achieved by using control loops to “measure” certain process parameters, “compare” them to specified set points (SPs) and then “adjust” the process accordingly.
- An alarm is incorporated into the system to prompt the operator to take action when the process runs out of control. Some people may ask why you should bother to install an alarm when there is a backup system, called an interlock, to deal with this situation. If they are properly trained, the operator will be able to make the wisest decision to override the control system and take corrective action. This will prevent the loss of production that would occur if drastic action were activated via the interlock system.
- SIS. If a process parameter goes out of control, an alarm is activated, which allows the plant operator to override the system to bring the process back under control. If he is unable to achieve this, the next layer of control comes into play. This is the safety instrumented system, whose main aim is to protect equipment.

Table 13.1 Manual versus automatic control.

	Manual control	Automatic control
Sunny days	Needed	BPCS
Rainy days	SIS “runs the show.” Manual control is not necessarily the best action.	SIS

At this stage, the SIS doesn’t care about process control. It employs direct action, which may halt the production process in order to protect the plant and its equipment. The action of a SIS system is also called a “trip.” The interlock system employs direct action that will most likely halt the production process. Its main function is to prevent equipment loss, safeguard plant personnel, and protect the environment.

- Mechanical relief. If the action by the SIS is not sufficient, the system moves to the last layer: relief. The purpose of relief is not only to protect the plant and its equipment, but also to protect the environment and ensure the safety of plant personnel. This is usually achieved by activating a pressure relief valve. You may wonder why there is no relief valve for any of the other process parameters like level, flow, and temperature. This is because an increase in any one of these parameters will lead to a corresponding increase in pressure in the plant and in the equipment. So, only a pressure relief valve is normally needed. This concept is covered in Chapter 12.

Here in this chapter we focus on the BPCS, alarm, and SIS.

13.3 Parameters Versus Steering/Protecting Components

The question that should be answered is whether protecting components should be placed for every single process parameter or not. The answer is no and process parameters should be selected for the selected ICSS components. We are going to answer this question regarding the BPCS and SIS. The alarm won’t be discussed here as it is an “acting” component.

Theoretically, each element of BPCS and SIS can be applied to each of the five process parameters: level, flow rate, pressure, temperature and composition. So you could have $2 \times 5 = 10$ steering/protecting loops for each piece of equipment (Figure 13.1).

There are two other items that may potentially increase the number of steering/protecting loops around a piece of equipment.

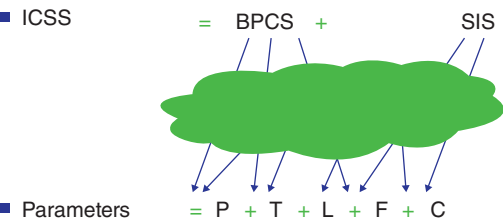


Figure 13.1 Swinging parameters versus steering/protecting components.

One is that in addition to process parameters there could also be non-process parameters and the other one is another steering/protecting loop for “only-monitoring” parameters.

The non-process parameters are not as common as process parameters. For example the torque of a rotating scraper at the bottom of a sedimentation basin is a non-process parameter. Non-process parameters, if used, can be utilized for monitoring or interlock purposes. More detail is provided in Chapter 16.

The monitoring parameters are the parameters that only need monitoring by the operator in the field or in the control room. This is different from the process parameters we need to control and, in addition to controlling them, we provide an opportunity for operators to monitor them.

Therefore there is a “minimum” of 10 different loops around each piece of equipment.

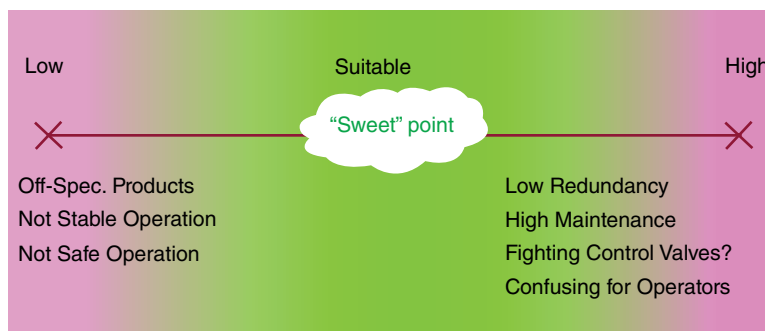
13.4 How Many Steering Loops Are Needed?

Placing control systems in a plant is a vital step to take during the design of any plant. But the question that needs to be answered is: how many control loops, alarms and SIS actions need to be implemented in a plant? There is no specific answer to this question. The responsible person needs to implement the “sufficient” number of control systems for each plant.

If the level of control is too low, you compromise the quality of the product. The operation could also be unsafe. The operation may lead to producing low-quality (off-spec) product. The operation could be very operator intense because of instability in the operation.

On the other hand, the larger the number of control loops, the more system maintenance is required. If the control loops are not well designed, they may compete with or counteract each other. Another thing to consider is that if you have installed a large number of alarms, this can lead to operator confusion because he is “overloaded.”

Figure 13.2 Level of control.



So, a designer has to find a happy medium between the two. The main thing to remember is not to over-complicate the control system.

This concept is shown in Figure 13.2.

13.5 ICSS System Technology

Until now, two main control “concepts” in process plants have been introduced: BPCS and SIS. However, each of these two concepts needs to be implemented in the plant through “hardware systems.” Basically, carriers are required in order to materialize these concepts in a plant.

There are currently two main technologies that are used as “carriers”: the so-called distributed control system (DCS), and the programmable logic controller (PLC).

Table 13.2 shows the three concepts of control: BPCS, alarm, and SIS and the respective hardware systems, which work as “carriers.”

As can be seen, the main role of the DCS is to handle BPCS actions, whereas the role of the PLC is to handle SIS actions.

To refresh your memory, SIS actions are on/off actions for the purpose of safety. These may be called “safety interlocks.”

The duty of the BPCS is to “adjust” the process to make it run smoothly. However, the action of the BPCS is not only regulating or throttling a valve (control valve); it could also be opening or closing a valve!

For example, in some semi-continuous operations you need to close some valves and open other valves to go from one cycle of operations to another.

One famous example is in sand filters, in water treatment plants. Sand filters cannot work purely continuously; they need to be pulled out of operation every so often to be backwashed, in order to remove trapped particles and to “rejuvenate” the filter for the next round of operation. This BPCS action is called a “discrete action” (as opposed to a regulatory action). Here we need to close and open some valves (switching valves); the intention is not safety but rather process. We call such actions “process interlocks.”

To summarize what we have said about control systems, for each component of the ICSS, we need to decide which technology to employ:

- **BPCS:** the control function of the BPCS can be either regulatory or discrete. DCS technology is most often chosen for regulatory control, through control loops. PLCs can be employed to control single unit operations within the process, or for batch control.
- **SIS:** we can define an SIS as a set of discrete functions for safety purposes. Here we need a very reliable system and it has been found that a PLC is more reliable than a DCS. For that reason, a PLC is usually selected for this role. However, this type is not a conventional PLC and is called a “safety PLC.”
- **Alarm:** could be run through a DCS or SIS

Now the question is: can we use a PLC for BPCS actions, or a DCS for SIS actions? This question is answered in the next two sections.

Table 13.2 Control system “carriers.”

	DCS	PLC
BPCS	The main BPCS actions are through the DCS	Some BPCS actions could be through the safety PLC
Alarm	Alarms through the DCS are acceptable	Alarms through the safety PLC are acceptable
SIS	No! Not good practice to run SIS through the DCS	All SIS actions are through the safety PLC

13.5.1 Use of PLC for a BPCS

The concept of using a DCS for a BPCS is not set in stone. There are at least two instances where it is possible, or even preferable, to use a PLC for BPCS actions:

- 1) **Batch operation.** A DCS is good for regulatory or adjustment actions. A PLC, on the other hand, is good for discrete control that does not require adjustment – this

is the case for batch operations, as was mentioned earlier. For example, in a filtration operation you have the filtration cycle, backwash, and then filtration again.

- 2) A very small plant. The DCS system was invented to control large plants, such as a huge refinery. It is not economical to install a DCS for a very small plant and in this case, a PLC is preferred for process control. Individual equipment is often supplied with a dedicated PLC from the manufacturer.

13.5.2 Use of DCS for a SIS

It is not good practice to use a DCS to handle SIS functions, because it is not nearly as reliable as a safety PLC.

13.5.3 Alarm Systems

The alarm system can be incorporated into the architecture of either the DCS or the PLC. There are some situations where alarms (or sometimes SIS actions) are incorporated neither into the DCS nor the PLC but rather are hardwired. There are at least two cases where it may be decided to hardwire an alarm (rather than incorporate it into the DCS or PLC).

The first case is if the alarm is extremely critical and you need to ensure that it is definitely activated, even in the event of failure of the DCS or PLC systems. Second, you could hardwire the alarm when the equipment is located in a remote area where no hardware (neither DCS nor PLC) is available.

We will talk about the symbology of alarms when we get to alarm systems at the end of this module.

13.5.4 ICSS System Symbology

As we mentioned, there is different hardware to handle the concepts of BPCS and SIS. When we are reading a P&ID, it should be clear whether the action is handled by the DCS system or the PLC system. Because of this, the symbologies are different for each case.

Table 13.3 shows the typical symbology used for DCS and PLC control systems.

A BPCS has a regulatory function via a DCS, and is represented by a circle inside a square. The way I like to memorize this is to equate the smooth control of a BPCS to the roundness of a circle.


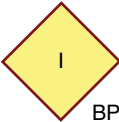


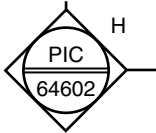




A BPCS may have a discrete function (a process interlock). This is represented by a diamond with the letter “I” for “interlock” inside, and the text “BPCS” on the outside to distinguish it from a safety interlock. It can also be shown as a diamond within a square with “BPCS” written outside, or alternatively as an octagon inside a square.

The main function of the SIS is as a safety interlock. This is represented by a diamond inside a square, either with the letter “I” or the letters “SI” inside it. It can also be shown as a diamond inside a square with the letter “I” in it and the description “SIS” outside. You can memorize the shape of the symbol by equating its harsh action to the sharpness and roughness of a diamond.

Using a DCS for a safety action is not good practice and there is no symbol for that.

Figure 13.3 shows an imaginary aerial view of a plant, showing a layer of regulatory control with a layer of safety interlocks above it. All of the safety interlocks are shown as diamond shapes to represent control by PLC,

Table 13.3 Example of control system symbology.

	Duty	DCS	PLC		
BPCS	Regulatory		 BPCS	 BPCS	
					
SIS	Safety interlock		 SIS	 SIS	 SIS

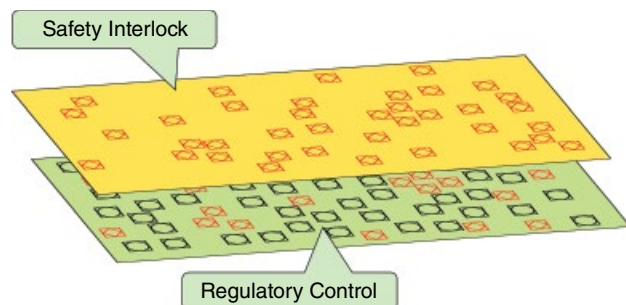


Figure 13.3 Imaginary layers of control.

because very few people use a DCS for this function. In the layer of regulatory control, the majority of shapes are circles to show control by DCS. There are diamond shapes here as well, to illustrate that some unit operations are controlled by PLC integrated with the DCS. These could be equipment packages bought from manufacturers with their own dedicated PLCs.

In the regulatory control layer, it becomes the responsibility of the I&C practitioner to integrate the PLC “islands” into the rest of the DCS.

13.6 ICSS Elements

Primary and final elements vary, depending on the type of action and function of the control system, as shown in Table 13.4.

(Primary and final elements will be discussed later but in a nutshell, the primary element is the sensor and the final element could be a valve or control on an electric motor)

The BPCS is discussed in this chapter and Chapters 14 and 15. In Chapter 16 alarms and SISs will be discussed.

13.7 Basic Process Control System (BPCS)

It is important to understand that BPCS control is not an add-on to a given process, but rather an integral part of it. A good understanding of the process is essential for a good control system.

The first question that needs to be answered is whether we need to implement a control system for a certain piece of equipment or not. Why should we bother to install a control system at all?

We have to install a control system for all elements, except for elements that meet the following three requirements:

- 1) Where there is no fluctuation. In the case of a fully closed circulating water system, for example, there is typically no fluctuation. An example of this would be a completely closed system, such as an HVAC system where the flow rate doesn't change.
- 2) Where fluctuation does exist, but it is not important (i.e. it is tolerable) in the process. Sometimes we might have fluctuation in a flow to a storage tank – but variations in flow to a storage tank are typically not considered important to a process. On the other hand, a fluctuation in flow to a reactor or a distillation tower would be a critical factor in a process.
- 3) Where the equipment is self-controlled. There are not many pieces of equipment that are self-controlled.

In chapter 14, various complicated loop systems are discussed, but we don't use these complex systems if we don't really need to. The best control system is a single-loop system.

Let's start our discussion about control loops with a simple daily life example of control.

Figure 13.4 shows Bill preparing to wash the dishes. The first thing he needs to do before filling the sink is to

Table 13.4 Control system elements.

	BPCS		SIS
Action	Regulatory	Process discrete	Safety discrete
Name	Control loop	Discrete control loop	SIF loop
Schematic			
Primary element	Sensor (sensitive toward a range)	Sensor or switch (sensitive toward a point)	
Final element	Control valve or VSD on electric motor	Switching valve or start-up/shutdown switch on electric motor	

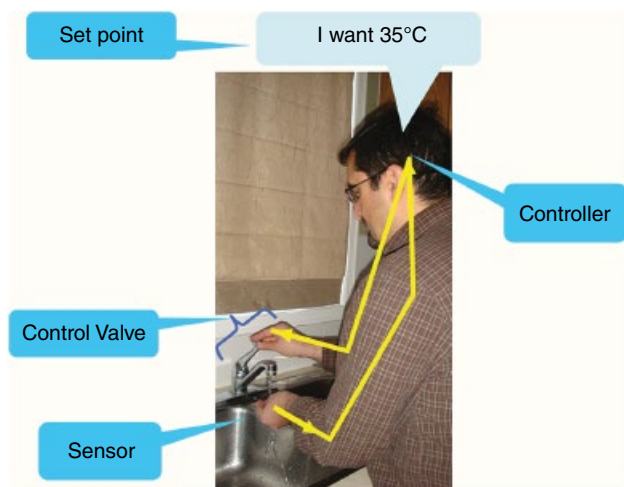


Figure 13.4 Everyday example of a control loop.

set the tap temperature. The following is the sequence of steps for doing this:

- 1) Bill's left hand (the "sensor") monitors the temperature of the water coming out of the tap.
The sensor sends a temperature signal to Bill's brain (the "controller"). The controller contains in its registry a "set point" of 35°C, a comfortable temperature for washing dishes.
- 2) The controller checks the reported temperature against its SP.
- 3) The controller sends a temperature adjustment signal to the control valve.
- 4) The control valve interprets the signal and adjusts the hot or cold water flow required to achieve a dishwater temperature of 35°C.

In this example, the "control valve" is the lever on the hot and cold water mixer tap. The "actuator" for the

control valve is Bill's right hand. The actuator adjusts the control valve upon receiving a signal from the controller, based on a SP that resides in the controller.

Now we want to show the above control loop with the symbology used in P&IDs (Figure 13.5).

We need to start with defining the devices involved in the control loop:

- TE: temperature element, or sensor
- TT: temperature transmitter
- TC: temperature controller
- TV: temperature control valve (the acronym is not TCV!).

Using the above definitions, we can outline the sequence depicted Figure 13.6. Note that it is assumed that the tap has been opened before step 1 of the sequence:

- 1) The TE records the tap water temperature.
- 2) The TT transmits this value to the TC. The transmission is represented by the dotted line.
- 3) The TC compares the received value against its registered SP.
- 4) The TC sends a temperature adjustment signal to the TV.
- 5) The TV adjusts the flow of hot and cold water.

All control loops include transmitters, whether they deal with temperature, pressure, level or flow. However, not all companies show transmitters on their P&IDs. The reason is that transmitters are essentially signal channels with no "process" duties, and some companies prefer not to show non-process items on their P&IDs. The purpose of the transmitter is to convert the signal created by the sensor into a signal that is "good," i.e. clear and strong enough to reach the controller, which is usually located in the control room.

All of the elements in one control loop need to have a single unique sequence number. In Figure 13.6, this

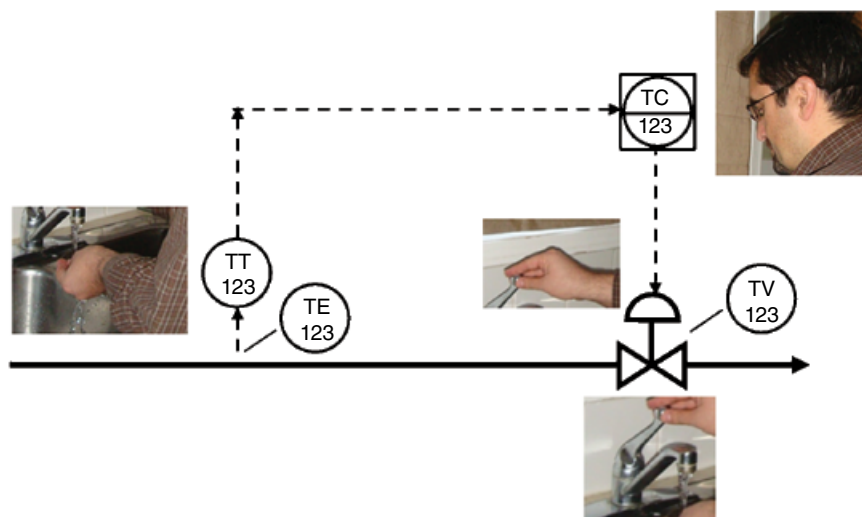


Figure 13.5 Temperature control loop.

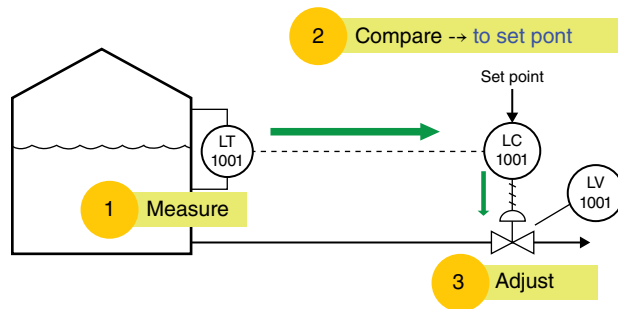


Figure 13.6 Level control loop.

number is 1001. Sometimes the elements of a single control loop do not all appear on the same P&ID sheet, so if these elements have all been assigned the same sequence number, we can still find them easily in different P&ID sheets.

Now, we will continue to look at the second example of a simple level control loop.

Figure 13.6 illustrates fluid level control in a tank, rather than fluid temperature control.

The devices in this loop are defined as follows:

- LE: level element, or sensor
- LT: level transmitter
- LC: level controller
- LV: level control valve.

There is no “LE” in Figure 13.7 and nothing is missed there! There are some cases that there is no specific sensor for a process parameter and the signal is initially developed in the transmitter, which is “LT” here.

For level measurement, there are two types of sensor: contact sensors and roof sensors.

Contact level sensors are the most famous sensors in that their signal is developed in their transmitter and no “LE” exists for them.

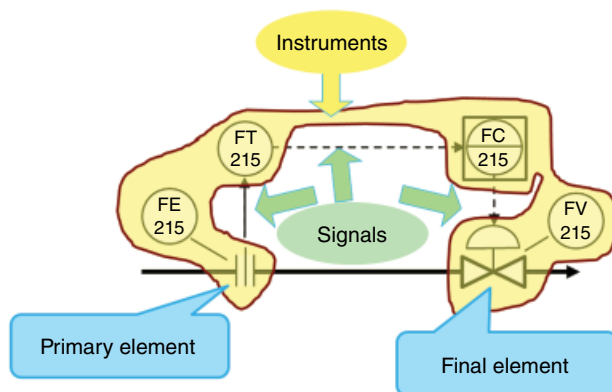


Figure 13.7 Fundamental terminology.

Using the above definitions, we can outline the sequence depicted in Figure 13.6:

- 1) The LT transmits the level value in the tank, as measured by the contact sensor, to the LC. The transmission is represented by the dashed line.
- 2) The LC compares the received value against its registered SP.
- 3) The LC sends a level adjustment signal to the LV.
- 4) The LV adjusts the control valve to modify the outflow from the tank in accordance with the instructions received from the LC.

For example, let us assume the SP to be 2m. This means the level needs to be constantly adjusted to 2m from the bottom of the tank. If the level is reported at 2.5 m, the LC compares this value against the 2m SP and recognizes that the tank contains an average of 0.5m more than the SP. Therefore, the LC sends a signal to the LV to adjust the level downward by opening the valve to increase the outflow from the tank.

This example illustrates that each control loop has three functions: measuring, comparing, and adjusting.

13.8 Instruments on P&IDs

A piping and instrumentation diagram (P&ID) is a schematic drawing used to illustrate all of the elements used in the control of a process. It is a diagram that shows how all the pieces of process equipment are interconnected, together with the instrumentation used to control the process. Symbols used for both equipment and instrumentation conform to the global guidelines given by the ISA, the International Society of Automation.

13.8.1 Fundamental Terminology

I would like to explain some fundamental terms that are used in process control, specifically relating to control loops (Figure 13.7). First, you have a primary element (sensor), which is usually an instrument that measures a process variable. A signal is sent via a transmitter to a controller. The controller then sends an adjustment signal to the final element.

The final element is some mechanical means to control the process. This is often a control valve on a pipe, or a variable speed drive for a pump or compressor.

13.8.2 Identifiers for Equipment and Instrumentation

Here we want to learn the identifiers of instruments. We will start our discussion with a table that compares equipment versus instrument identifiers.

Table 13.5 Instrument identifiers.

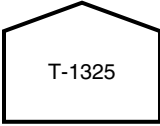




	Equipment	Instruments
Example	T-1325 2.2 KPag @80C 	
Call-out		
Tag		Two to four (or more) letter and tag number
Symbol	Different for different equipment	For sensors: different symbols for different Sensors For the rest of instruments: different balloon shapes
Symbol represent:	Actual shape of equipment	Function of instrument and the location of instrument

Table 13.5 gives examples of the typical identifiers that are used for both equipment and instruments. The identifiers used for equipment are different to those used for instrumentation.

For equipment, we use the actual shape as a symbol; there is a different one for each type of equipment. Each piece of equipment has an identifying tag and a call-out, which gives key information about the operating parameters of the equipment. The call-out information can be inserted at the top or the bottom of the P&ID.

On the other hand, instrument symbols don't resemble the actual shape of the instruments. They specify the function and the location of the instrument. Instrument symbols (or "balloons") generally conform to ISA specifications.

For instruments, we do not use a call-out. A tag is used to describe the function of the instrument and is between two and four letters long.

13.9 Instrument Identifiers

When we show an instrument on a P&ID, there are five items that we need to specify (Figure 13.8):

- 1) Acronym
- 2) Divider type
- 3) Symbol type
- 4) Additional information
- 5) Tag number.

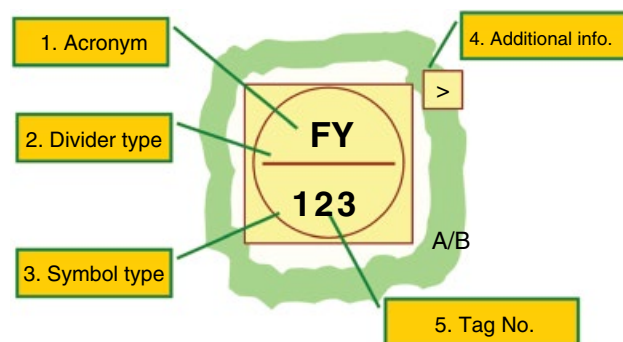


Figure 13.8 Instrument identifiers.

We will discuss all of the identifiers except the last one (tag number). The tag number is simply a sequence number for an instrument. A big help when reading a P&ID is that all the elements in one control loop have the same tag number. This fact helps us to find different elements of one control loop when they are on different P&ID sheets.

13.9.1 Acronyms

Acronyms are composed of two "bags"; one left hand bag and one right hand bag, as shown in the schematic in Figure 13.9. Each "bag" may have one or two letters. The second letter of each "bag" is not mandatory. Therefore, we can say that an instrument's acronym can have between one and four letters.

The meaning of each letter in the tag's acronym is presented in Figure 13.9.

The first letter is mandatory and is used to describe the process parameter that is being controlled. Most of the first letters of the acronym are self-explanatory, except "A." "A" represents composition, based on the word "analyzer," which is the word used by chemists and lab technicians for any composition parameter being measured.

The second letter is called a "process parameter modifier." This is an optional parameter, and we use it

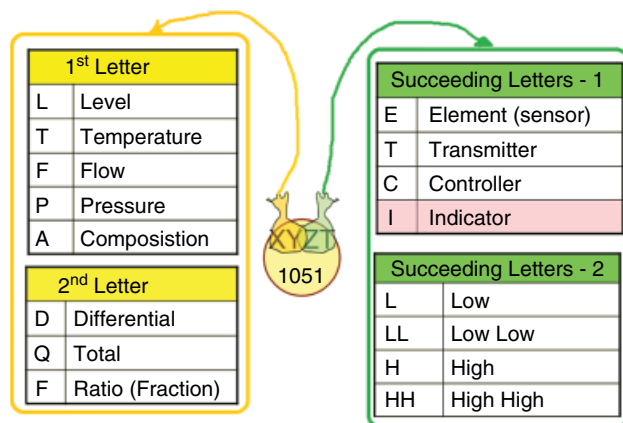


Figure 13.9 Acronym descriptors.

when the process parameter of interest is not a simple parameter. In other words, it involves some computational aspect of the parameter, e.g. ratio, total or differential.

For instance, “PD” would indicate the measurement of a pressure differential between two points in the process. The use of PD is of great benefit wherever flow goes through an obstructed route like in filters, strainers or even ion exchangers. The PD is used to ascertain when the porous medium has become plugged and needs backwashing. The modifier of “D” is widely used for pressure but it is not common for other parameters.

Another example is the use of “Q.” “F” means flow rate, but “FQ” means volume (total of flow rates means volume!). The modifier of “Q” is used arguable only for flow rate as it is not generally meaningful for other parameters. The total flow rate could be important for some utility streams, on some intermittent flow pipes, or other cases.

The use of the ratio descriptor is very useful for flow, and the acronym for this would read “FF.” This is of particular importance in dosing systems, where you need to control the dosing rate according to the flow rate of the process stream.

The third letter is called the function letter, which is what we want to do with the parameter obtained and mentioned in the parameter of choice stated in the first letter. If the instrument is sensor, we use the letter “E”(element); if it is a controller, we use “C,” etc.

If the instrument is an indicator, we use the letter “I” but usage of “I” is not as common as in older days. The reason will be discussed in Section 13.9.3.

The fourth letter in the acronym is, again, an optional descriptor. The fourth letter is a modifier that is most often used for alarm purposes or in SIS actions to indicate the action point of a loop, for example, high, low or low-low.

Table 13.6 gives more examples of instrument acronyms.

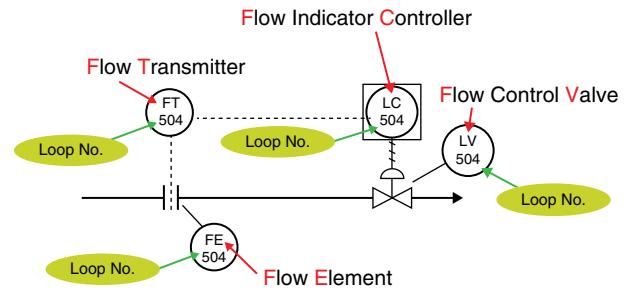


Figure 13.10 Instrument acronyms shown on a P&ID.

Figure 13.10 gives one complete example of a control loop.

13.9.2 Divider Types

The balloon dividers generally specify the “location” of instruments. What is important from an I&C practitioner’s viewpoint regarding “location” is if the instrument is in the field, in the control room, or in the field cabinets.

Different divider shapes are shown in Table 13.7.

In Table 13.7, I have shown the symbol with an irregular shape. This is so that you focus on the divider and not the shape. The shape of the symbol will be discussed in Section 13.9.3.

For the divider, there are five different cases:

- 1) No divider. This means that the instrument is outdoors, in the field. Examples are a flow element or sensor, or a level switch or gauge. They are connected to the control system, but they are not encased in a control room or in an auxiliary control cabinet. The majority of sensors are located outdoors and their tag doesn’t have any divider.
- 2) Single solid line _____. This shows that the instrument is situated inside the main control room. It also indicates that the instrument is accessible and visible to the operator.

Table 13.6 Examples of instrument acronyms.

No.	Examples	What is the parameter we are looking for?		What we want to do with the parameter?		Meaning
		Parameter of interest	Parameter modifier	Function	Function modifier	
Example 1	FQI	F	Q	C		Volume of fluid is controlled
Example 2	LT	L		T		Liquid level is transmitted
Example 3	TC	T		C		Temperature is controlled
Example 4	PDC	P	D	I		Pressure difference is shown
Example 5	AT	A		T		An analyte is transmitted
Example 6	LEHH	L		E	HH	Level sensor alarms on “HH” level

Table 13.7 Divider types.

Location	Unenclosed location		Enclosed locations		
	In field	In accessible location	In control room	In field cabinet	
			In inaccessible location	In accessible location	In inaccessible location
Symbol (divider type)					
Examples	All sensors, all transmitters	Controllers, indicators	Non-important functions	Local alarms	Transducers

- 3) Single dashed line . This shows that the instrument is located inside the main control room, but it is not that important and is inaccessible to the operator. In the past, before computerized control, instruments were mounted on a board in the control room. The important ones were mounted on the front of the board and those that were deemed less important were mounted on the back of the board. So, this is the same concept.
- 4) Double solid line . This indicates that the instrument is situated in a control cabinet in the field and is accessible and visible to the operator. Dedicated control cabinets are supplied by manufacturers, especially for PLC control of their equipment.
- 5) Double dashed line . This means that the instrument is in a field control cabinet, but inaccessible to the operator.

The way that you can memorize these cases is this statement:
“no divider; no enclosure,” the rest are in enclosures.

The meanings of different dividers are summarized in Figure 13.11.

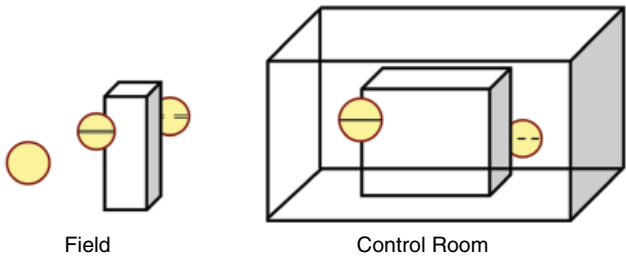


Figure 13.11 Instrument divider types versus their location in reality.

Why do we use dashed lines?
In this symbology, we use dashed lines based on the interpretation we borrowed from drafters. Did you know that drafters draw hidden lines as dashed lines? We use the same concept here. The items are behind a board are invisible and we show them with a dashed line.




13.9.3 Symbol Type

We talked about this previously in this chapter. There are three different types of symbols for instruments, as shown in Figure 13.8.

The three types you see represented are:




- 1) A circle on its own. This represents an instrument that gives a measurement or readout. It has no controlling function, i.e. it has no brain.
- 2) A circle within a square. The square shows that the instrument has some controlling function. The circle represents a smooth control process, such as a DCS.
- 3) A diamond within a square. Once again, this instrument has a controlling function. The diamond indicates that the control process is via a tougher system like a PLC. This type is also sometimes shown as an octagon within a square.

The way that I memorize this is:
A single symbol means “no brain.” A double symbol means “a brain” is involved.
This has a circle with a soft and round perimeter, similar to the soft DCS system.
This has a diamond with sharp, tough corners, similar to the tough SIS system.

It is important to note that we show symbols like  or  not only when they have a “brain” (controller), but also when a piece of data can be seen through the monitor connected to the control system. This justifies symbols like this ; it is not a brain, it is only an indicator, but we use a circle to show it is visible through a DCS monitor.

By combination of Tables 13.7 and 13.8 we can see the famous table of instrument symbology that you may

Table 13.8 Types of instrument symbols.

Symbol (balloon type)	Non-technical definition	Example
	Whatever has no brain	Sensors, transmitter
	Brain in DCS and whatever can be “seen” in DCS monitors	Whatever other than above! But not part of SIS
	Brain in PLC	The logic holder in SIS loop






















have seen several times on the first sheet of a set of P&IDs (Table 13.9).

Table 13.9 is a table that was originally developed by the ISA and is well known in the control and instrumentation industry. It shows the types of symbols used for instruments in different locations.

With current technology there are some symbols that are no longer common. The reasons are as below:

- Field-mounted items: “discrete instrument” means an instrument with no controlling function. These are basically sensors and transmitters. As we know, sensors are installed on processes, which are located in the field (i.e. not inside the control room or a field cabinet). Transmitters are connected to sensors and as close as possible to them. That is why they generally don’t have any divider for sensor and transmitter symbols. It doesn’t make sense to install a sensor or transmitter inside a control room or inside a field cabinet. Therefore, it is rare to see any type of divider for a sensor or transmitter.
- “Brains,” either DCS or PLC, are too delicate to be installed in the harsh environment of the field. Therefore, there is no “square”-type balloon, with no divider. DCS and PLC systems are generally installed indoors, so they have some sort of divider.

Table 13.9 Instrument symbology.

Non-technical definition	Unenclosed location	Enclosed locations			
	In field	In control room		In field cabinet	
		In accessible location	In inaccessible location	In accessible location	In inaccessible location
Whatever has no brain	 				
Brain in DCS and whatever can be “seen” in DCS monitors		 		 	
Brain in PLC		 		 	 

- Back-of-panel items: there is no such thing as “back-of-panel” in today’s control rooms. In the old days there was a real panel in control rooms and some not very important instruments could be checked and monitored only by going to the back of the panel. However, these days all the panels in control rooms have been replaced with monitors, thus there is no panel of a back of panel. So, a single dashed line, — — —, is very rare.

However, we still have backs-of-panels for field cabinets, i.e. = = = =.

13.9.4 Additional Information and Tag Number

There are no set rules as to where to put additional information about the instrument on a P&ID. However, usually you place it at the top right or bottom, right next to the symbol, as shown in Figure 13.12.

There are at least three different instances where you need additional information for the instrument in question:

- 1) For function clarification
 - 2) For carrier clarification
 - 3) For tag clarification.
 - Function clarification. Different examples are shown in Figure 13.12(a). Generally speaking, however, wherever there is a non-defined letter in an instrument symbol, it needs to be specified outside of the symbol.
- To explain further the concept of a “non-defined” letter we need to refer to table 4.1 of ISA standard ANSI/ISA-5.1. The non-defined letter could be a “user’s choice” letter (like “M”), or an “unclassified letter (like X)”, or letters that refer to general actions (like “Y” or “U”), which are not precisely defined.
- In the first example we see “FY”; “Y” is a mathematical operator that is applied to a flow, “F”

The question is, “what is Y?” We have specified it outside and put the symbol “>,” which means “larger than.” This means that this balloon receives more than one flow signal, and this balloon will pick the largest flow rate.

Another example is a UA balloon. “A” is alarm but “U” is a wildcard letter and needs to be specified; outside of the symbol it says that “U” refers to any problem in the system, or “common alarm.”

The last two examples are interesting because the letters used in symbols are clearly defined by the P&ID developer deciding to “clarify” them outside of the symbols.

Another example is PI. This example is different than the previous ones. Here there is no wildcard or unknown parameter. However, some people could be confused about the purpose of measuring and indicating pressure in that specific application. In that case, the designer would explain that “we are measuring the pressure to make sure there is no leak from the system.”

Yet another example (not shown here) is a composition element, AE. In this case, you need clarification by showing what analysis is being conducted, e.g. pH or H₂S in gas, because “A” doesn’t tell us what the analyte is.

- Carrier clarification. As shown in Figure 13.12(b), we can clarify the carrier, or type of control system, that the instrument relates to. This is written on the bottom right-hand side of the tag and can be “BPCS,” “SIS” or “BMS.”

A “BMS” is a “burner management system.” This is essentially an SIS that is a critical and mandatory part of burner management in industry. In some jurisdictions BMS is mandatory when you deal with a fired system.

- Tag clarification. In Figure 13.12(c) we show an example of a tag for a flow controller, FC, which is the same for two different items, A and B. To clarify this, we show the lettering A/B on the side of the symbol.

The other example shows an interlock, or “I.” This has the number 1091 written on the side. The number 1091 relates to the action that would be taken by the SIS. You would have to refer to the cause and effect diagram, or shutdown key, to see what action would be taken when SIS loop 1091 is activated.

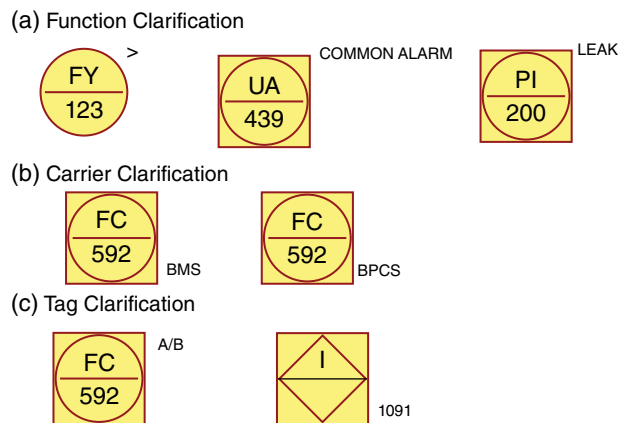


Figure 13.12 Additional information.

13.10 Signals: Communication Between Instruments

Communication between control instruments is via different types of signals. We need to have a look at how this is shown on P&IDs (Figure 13.13).

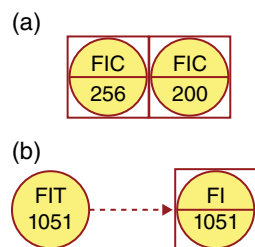





Figure 13.13 How do “boxes” communicate?




If the two elements are situated within one housing, we show them side by side with no evidence of a signal (Figure 13.13(a)). In reality, you need something to carry a signal between the two but this is obviously part of the internal wiring of the particular control panel. If the two instruments are situated in remote locations, you need to show a signal line connecting them (Figure 13.13(b)).

13.10.1 Signal Types

There are different types of signals that are used for communication in process control instrumentation (Table 13.10).

- Capillary line . A capillary line is a tube with trapped innocent liquid. This tube is used when we try to “transfer” the process fluid to an instrument without sending the process fluid inside of the instrument, possibly because the process fluid is non-innocent. This is primarily used for sensors in some special cases.
- Electrical signal . An electrical signal can be either analog or digital. This has traditionally been an analog signal, but the latest technology is moving more toward the digital form. Although there is a







difference in an analog-electrical signal versus a digital-electrical signal, we can consider  as the general symbol for electrical signals, whether analog or digital. Electric signals can be considered as the “default” signal in today’s process plants.

- Pneumatic signal . The majority of control valves are actuated by pneumatic signals. The pneumatic signal is generally provided by “instrument air” (IA). However, other gases could be used if IA is not available. This type of signal is used for the majority of control valves and switching valves.
- Hydraulic signal . In some cases, hydraulic actuation is necessary for the operation of a control valve or switching valve. First, in a remote location like an oilrig, an instrument air facility may not be available. Second, if the pressure in a pipe is very high and the instrument air pressure will not be able to adjust the valve, then hydraulic oil is needed to open and close the valve. In such cases, a hydraulic oil package needs to be purchased, instead of an instrument air package.
- Software signal . Software signals are actually computerized signals. The simplest form of a software signal may be established by an RS-232 cable, which is very famous in the computer industry. In plain English, a software signal is a digital signal whereby digits are sent in a specific packet size and a specific time/duration. These signals are sent where two controllers (balloons with “C” in them) want to communicate with each other.

There are some companies who put arrows on their signals and there are other companies that don’t follow this practice. In either case, to correctly interpret a control loop, the P&ID reader should put arrows on the signals in his mind, even where they don’t exist.

Now the question is what we need where a signal extends beyond a P&ID sheet. The answer comes with the same logic of pipe extension: off-page connectors. Figure 13.14 shows off-page connectors for a signal.

Table 13.10 Types of signals.

Signal symbol	Signal name	Common application
	Capillary tube	To some fluid-in sensors
	Electrical-analog signal	By default choice
	Electrical-digital signal	Advanced choice
	Pneumatic signal	On pneumatic control valves that are very common
	Hydraulic signal	On hydraulic control valves
	Software link	Connection between “brains”

13.10.2 Signal Functions

Signal functions are drawn as balloons that process signals to convert them to other types of signals. There are two main categories of signal functions: signal math

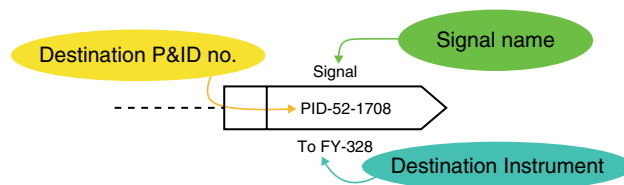


Figure 13.14 A leaving off-page connector for a signal.

functions and signal selectors. We will explore these two in the following two subsections.

13.10.3 Signal Math Functions

Signals are not always straightforward and sometimes we want to add them, multiply them, divide them, and also use the difference between them to create a new control signal.

There are control applications where we need to use a function curve, $f(x)$. For example, when you need to control the pH of a stream/liquid inside a vessel, the control will be based on a titration curve. The titration curve is digitized to form a function curve, which is then used to control a neutralization system.

Some of the most common math functions are listed in Table 13.11.

13.10.4 Signal Selectors

In some instances, we have more than one signal available for control, and we need to select one of them. Table 13.12 shows the various options that are most often used for signal selection. In the next chapter you will see some application of these signal selectors.

Figure 13.15 gives examples of how signal selectors work. These are quite self-explanatory for each case. Limiting selectors have an internal “constraint.”

Table 13.11 Signal math functions.

Signal symbol	Signal name	Meaning
<div><div>Σ</div><div>+</div></div>	Summing	Adding two signals together
<div><div>Δ</div><div>−</div></div>	Subtracting	Subtracting one signal from other signal
<div><div>±</div></div>	Bias	Subtracting one signal from a constant value
<div><div>x</div></div>	Multiplying	Multiplying two signals together
<div><div>k</div></div>	Proportional	Multiplying one signal to a constant value
<div><div>÷</div></div>	Dividing	Dividing two signals
<div><div>Σ/n</div></div>	Averaging	Averaging several signals
<div><div>√</div></div>	Root extraction	Measuring the root of a signal
<div><div>f(x)</div></div>	Function	Applying a complex function on a signal
<div><div>d/dt</div></div>	Rate	Measuring the rate of a signal

Table 13.12 Signal selectors.

Symbol	Selector name	Function
<div><div>></div></div>	High selector	Picks the higher magnitude signal
<div><div><</div></div>	Low selector	Picks the lower magnitude signal
<div><div>⤴</div></div>	High limiter	Picks the higher magnitude signal up to certain value pre-set for it
<div><div>⤵</div></div>	Low limiter	Picks the lower magnitude signal down to certain value pre-set for it
<div><div>HS</div></div>	(Soft) hand switch	Picks a signal based on the operator intent

It is good to know that a “magnitude limiter” is the same as a “magnitude selector” when the magnitude selector function is constrained with a constant value rather than another signal (Figure 13.16).

You might assume that in order to make a selection, the signals need to be homogenous, i.e. both must come from a flow sensor or temperature sensor, etc. This is not always the case; the signal selector does not care where the signals come from – they only look at the value assigned to each signal in order to make a selection. Figure 13.17 gives an example of a non-homogenous signal selection scenario.

In the cases of non-homogeneous selectors the problem is how to tag the selector. If two signals are flow signals the selector can easily be tagged as “FY” and if

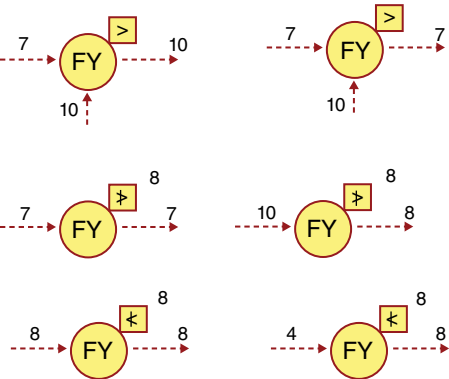


Figure 13.15 Examples of signal selection.



Figure 13.16 Magnitude limiter and magnitude selector.

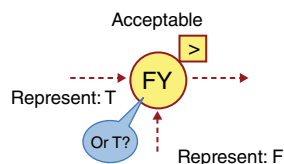


Figure 13.17 Signal selector on non-homogeneous signals.

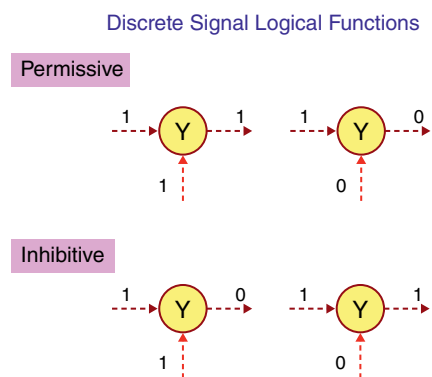


Figure 13.18 Discrete signal logic functions.

two signals are temperature signals the selector can easily be tagged as “TY.” Some companies drop the first letter and some other companies only show the higher or lower sign instead of showing the whole symbol.

There are some signal selectors that are specifically used for discrete signals. These are mainly permissive operators, and inhibitive operators.

The meaning of them is shown in Figure 13.18. Although I show the meaning of these two functions schematically, on P&IDs generally they are not shown. The permissive and inhibitive functions are generally captured on P&IDs only by notes.

- **Permissive.** The selector will only send an outgoing signal if both incoming signals are present, i.e. assigned a “1.” If one of the signals is absent (a “0”), then the selector does not forward a signal. This function is used in applications where one action needs to be performed in order to activate the other action. For example, an operator pushbutton for shutting down a centrifuge may have a permissive function of the open wash water valve. In this example, pushing the shut-down button of the centrifuge doesn’t work unless the system is already being washed by water through the opening of the wash water stream.
- **Inhibitive.** This selector works in the opposite way to the permissive selector. This means that if there are signals from both incoming sources, the selector will not produce an outgoing signal. Only if either of the incoming signals is absent will the selector send a signal.

13.11 Different Instrument Elements

In this section the main elements of a control loop are discussed. They are primary elements (sensors), transmitters, controllers, indicators, and the final elements.

13.11.1 Primary Instruments

There are three main types of primary elements. They are sensors, gauges, and “gauge points.”

Sensors and gauges, both, have sensing components and both of them are used to measure process parameters, but the former generates a signal (e.g. electric signal) proportional to the magnitude of the process parameter while the latter does not. This shows their applications; gauges are installed when the intent is to show a process parameter in the field to the rounding operator, whereas an element’s reading can be seen in the field and in the control room. Therefore a field gauge cannot be part of the control system.

“Gauge points” are nothing except a “point” that is “planted” somewhere on the plant to be used by an operator to check the process (or non-process) parameter with his portable gauge.

From this, you can deduce that gauges and sensor point never ever connected to other instruments on P&IDs.

All the primary instruments are installed in field then they don’t have any divider in their symbols.

Table 13.13 shows a list of gauges.

Elements have a more complicated structure than gauges. Gauges are not very complicated. For example, a temperature gauge (TG) could be a bimetal, wherein dissimilar expansion of an attached metallic ribbon causes its deviation and movement.

FGs are nothing more than a piece of transparent pipe that can be installed along a portion of piping. FGs are named “flow sight glasses.” The FG gives the operator a visual indication of flow.

The P&ID symbol for a sight glass can be as shown in Figure 13.19.

Table 13.13 Gauges.

Parameter	T	P	L	F	A
Tag	TG	PG	LG	FG	Not available

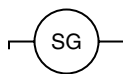


Figure 13.19 Flow sight glass.

Measuring process parameters is uniquely done by “process analyzers,” or “AEs.” They are too complicated to be built in the form of AG or gauge.

However there are some gauges than can be connected to indicators too.

Table 13.14 shows a list of sensors.

Some companies use acronym for gauges instead of elements interchangeably.

The portable gauges need specific locations on the plant to be able to “cling” on to them to measure the parameters. These “sensor locations” should be tagged on the P&IDs. A list of them is given in Table 13.15.

It can be seen from the table that gauge points are generally limited to pressure points and temperature points, or thermowells.

Some companies, instead of the acronym “PP”, use “PT”, which means “pressure tap.”

Now we are going to give a brief review of primary elements or sensors. It is important to stress that this is only a “summary” for the purpose of P&ID development. There is a huge amount of knowledge regarding sensors and the selection of them for different applications is beyond the scope of this book.

In this section we talk about each of five sensors, temperature sensors, pressure sensors, level sensors, flow

meters, and process analyzers. For each of them we briefly talk about each group and the common types of sensors. Then the different methods of connecting sensors to process items are shown. These methods are known as “hook up” arrangements. The hook up arrangements are not always shown on P&IDs, but it doesn’t mean they don’t exist. Some companies prefer to not showing them just to avoid crowdedness on P&IDs.

The P&ID symbol for each of the sensors is introduced here too. However, sometimes, especially in the early stages of P&ID development, the exact type of a sensor is not yet decided. In such cases just the tag of the sensor may be used as its symbol. Obviously this should be replaced later with the exact symbol of the selected sensor.

13.11.1.1 Temperature Measurement

Temperature measurement can be done everywhere on flow of gas, liquid, or flowable solids.

It also can be done for parts of equipment. One example is “skin temperature,” which is the temperature of the coil wall inside fired heaters. The other example is measuring the temperature of winding in high power electric motors.

A unique feature of temperature measurement is that they are types of sensors that work remotely. For example the temperature a flue gas stream can be measured remotely.

Temperature sensors generally don’t have any specific symbol on P&IDs.

Table 13.16 is a non-exhaustive list of common temperature sensors.

Table 13.14 Sensors.

Parameter	T	P	L	F	A
Tag	TE	PE	LE	FE	AE

Table 13.15 Gauge points.

Parameter	T	P	L	F	A
Tag	TW	PP	Not generally needed	Not generally needed	Not available

Table 13.16 Temperature sensors.

Type	Unique advantage	Unique disadvantage	Application
Thermocouple	Default choice for wide temperature range	Sensitive to noise	Specially in burner management system (refer to Chapter 16)
Resistance temperature detectors (RTD)	Very accurate	Needs power source	Process pipeline temperature measurement
Thermistor	High accuracy	<ul style="list-style-type: none"> • The narrow range • The most inexpensive temperature sensor 	More in laboratory but not in industrial process control.
Infrared	Non-contact type	Needs line of sight	Common for remote temperature sensing in combustion systems

Temperature sensor arrangements: when the pipe size is small it is not easy to place a thermocouple in it. “Small” is defined differently in different companies. Some companies consider small to be sizes smaller than 4” but some other companies consider 3” or even 6”.

There are two solutions when faced with the issue of placing a thermocouple in small size pipes. One is using a combination of enlarger–reducer to enlarge the pipe size and then placing the thermocouple (Figure 13.20).

The other solution is placing the thermocouple on a bend of the pipe (Figure 13.21).

If the purpose of temperature measurement is skin temperature, it could be mentioned beside the balloon on the P&ID, or showing the connection to the surface. Figure 13.22 shows a P&ID with both of them together.

13.11.1.2 Pressure Measurement

Pressure measurement can be done everywhere on the flow of gases or liquids.

However, measuring the pressure of liquids is not as common as gases and vapors.

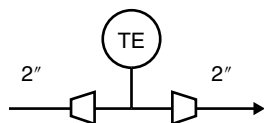


Figure 13.20 Thermocouple installation on narrow pipes (option 1).

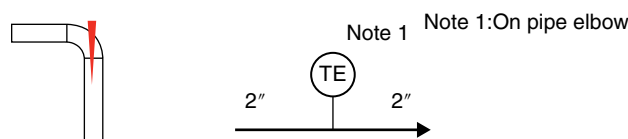


Figure 13.21 Thermocouple installation on narrow pipes (option 2).

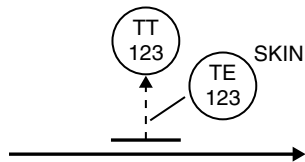


Figure 13.22 Skin temperature sensor.

Temperature sensors generally don't have any specific symbol for P&IDs.

Table 13.17 is a non-exhaustive list of common pressure sensors.

Pressure sensor arrangements: some companies show PGs and PIs simply as in Figure 13.23.

However, this doesn't necessarily mean that there is nothing between the gauge and the process. There are generally different items on the connecting tube.

As a minimum there should be one isolating valve – a root valve – and a drain/calibration valve (Figure 13.24).

In some critical cases, instead of one root valve, two root valves and a drain can be placed (Figure 13.25).

In the cases where the process fluid is harmful to the internal parts of the sensor, a diaphragm can be placed and the space between the diaphragm and the sensor

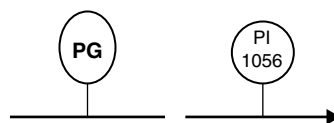


Figure 13.23 Pressure gauge and pressure indicator connected to process.

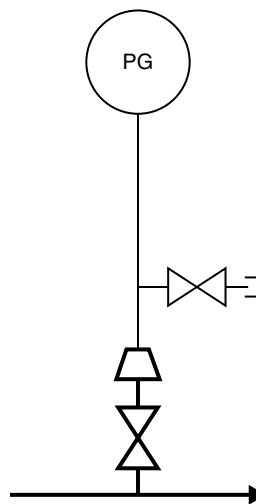


Figure 13.24 Pressure gauge with root valve and drain/calibration valve.

Table 13.17 Pressure sensors.

Type	Unique advantage	Unique disadvantage	Application
<ul style="list-style-type: none"> • Bellows type • Diaphragm type • Bourdon tube type 	Simple system	More prone to mechanical failure	Default choice
Piezoelectric	No moving parts	Limited range	General process pressure measurement

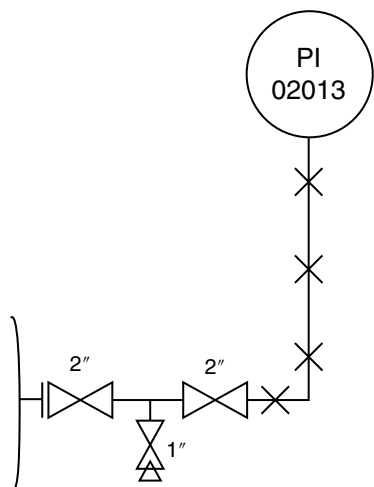


Figure 13.25 Pressure gauge with double block and bleed arrangement.

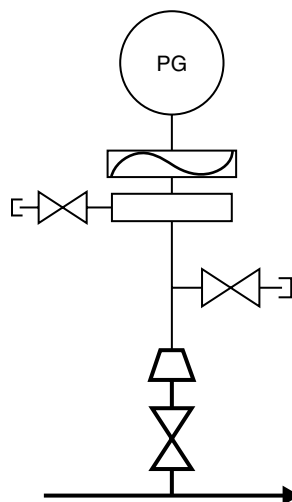


Figure 13.27 Pressure gauge with flush ring.

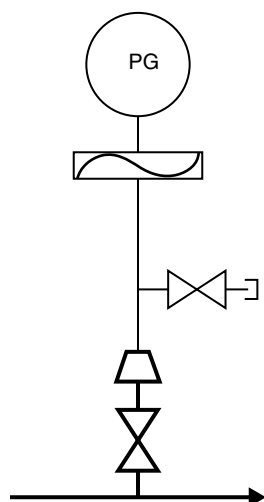


Figure 13.26 Pressure gauge with diaphragm.

filled with a suitable liquid, like glycol. An elastomeric diaphragm prevents the aggressive fluid getting into the sensor, while transferring the pressure (Figure 13.26).

This is a good practice when dealing with process fluids like precipitating liquids, corrosive fluids, and dirty fluids.

There are cases where the process fluid is very dirty or very precipitating that may leave sturdy settlement, preventing pressure transfer. In such cases a flush ring (bleeder ring) can be placed to clean out the connecting tube from time to time (Figure 13.27).

13.11.1.3 Level Measurement

Level measurement is something only for liquids or flowable solids in containers (vessels, tanks, basins, silos, open channels).

Level sensors can be categorized into two main types: contact type and non-contact type. Non-contact type level sensors are very attractive options where the liquid is aggressive and also they are the only choices for sensing flowable solid levels.

There are some types of level gauges that can be connected to indicators too.

Table 13.18 is a non-exhaustive list of common level sensors.

Level sensor arrangements: the connection of static pressure type level gauges and level transmitters is the same as for pressure gauges discussed before. The connection could be simply by a single root valve or through a double block and bleed arrangement. A diaphragm can be used if the liquid is aggressive. In such cases the process gets to the level transmitter through capillaries (Figures 13.28).

The process fluid gets to the level transmitter through capillaries (Figure 13.28).

The level gauges are connected to the containers through narrow pipes but not capillaries. The vertical portion of the connecting pipe is always equipped with a drain valve and vent valve (Figure 13.29).

In Figure 13.29 a level gauge is shown that is also connected to a level transmitter. This is one of the non-common cases in which a transmitter can be connected to a gauge.

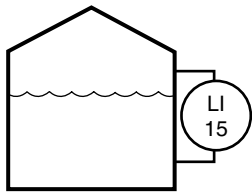
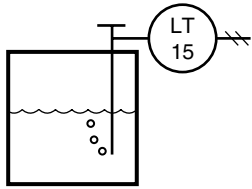
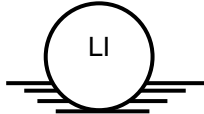
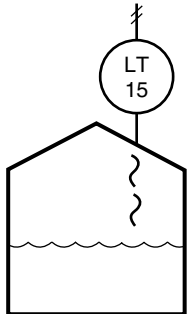
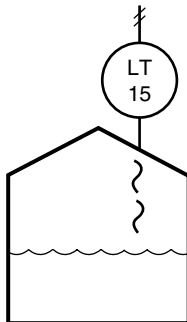
13.11.1.4 Flow Measurement

Flow measurement is done for liquids, gases/vapors and combination of them as multiphase flow.

Flow measurement of multiphase flows is more difficult and there are fewer options to choose for the measurement.

Liquid flow measurement can be needed in sealed flows or non-sealed flows. Sealed flows are the flows that fill the whole cross section of the liquid conductor while

Table 13.18 Level sensors.

	Type	P&ID schematic	Unique advantage	Unique disadvantage	Application
Contact type	Static pressure type		Simple system	Relays on density of the liquid that could be changing	By default choice
	Bubbler type		Good choice for slurry and precipitating liquids	<ul style="list-style-type: none"> Needs utility air connection Relies on density of the liquid 	Slurry liquids, water tanks
	Float type		Simple to operate	Limited range	Small tanks
Non-Contact type	Ultrasonic type		<ul style="list-style-type: none"> No contact with the process material 	<ul style="list-style-type: none"> The atmosphere should be transparent free of dust and liquid drops The atmosphere should be with fairly constant composition Liquid surface should be free of ripples and foams 	<ul style="list-style-type: none"> Relatively inexpensive
	Radar type (microwave)		<ul style="list-style-type: none"> No contact with the process material Elevation of interface in multi-layered fluids (oily water, water and sludge) can be measured. The measurement is not affected by the atmosphere condition 	<ul style="list-style-type: none"> More expensive 	Last resort

non-sealed flows are the flows where liquid is in a portion of the liquid conductor. For example the flow in open channels is a always non-sealed type conductor.

Table 13.19 gives a non-exhaustive list of common flow meters for sealed conductors.

Liquid flow in non-sealed conductors also needs to be measured. The fundamental principle of measuring liquid flow in open channels is to measure the liquid level, and then convert the level to the corresponding flow rate. Therefore the flow sensors in open channels are nothing other than level sensors.

Table 13.20 shows two types of open channel flow sensor.

Flow meter arrangements: there are some cases that a flow meter needs a fluid velocity higher than the pipe flow velocity to be able to sense the flow. In such cases the flow meter cannot be placed directly on the pipe. The pipe size should be shrunk to a smaller size (possibly one or two sizes smaller than the pipe size) and then the flow meter can be installed correctly. To do that a combination of a reducer–enlarger can be used (Figure 13.30).

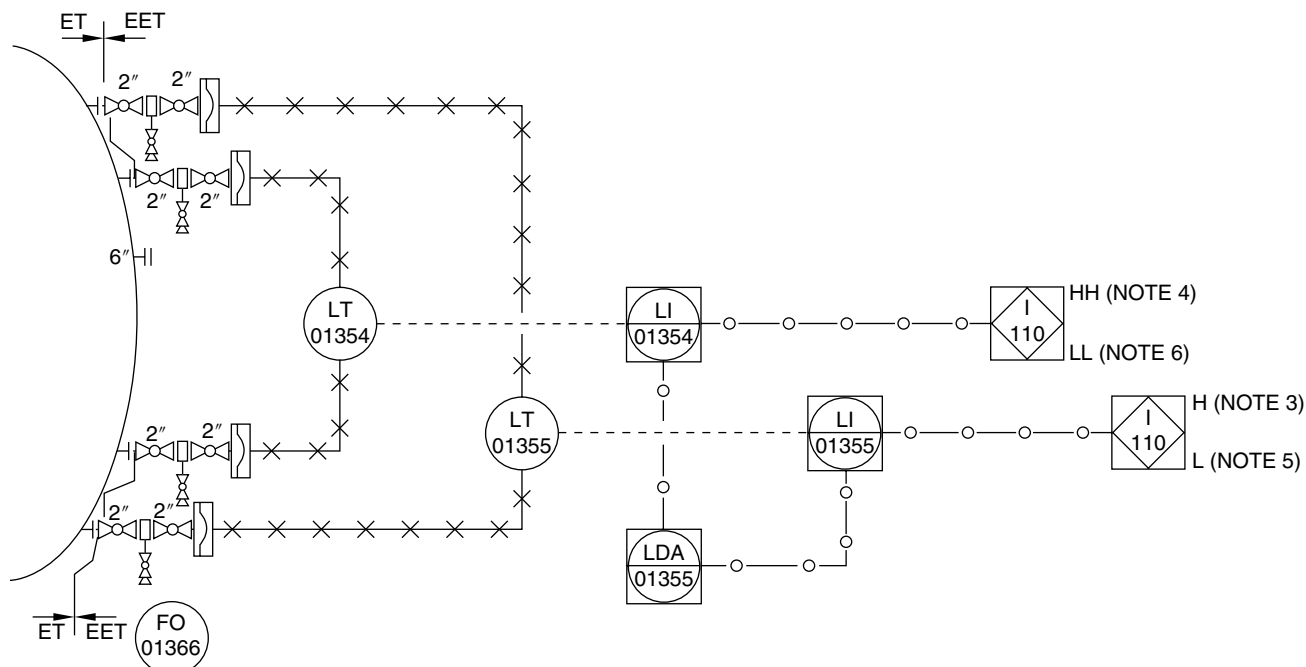


Figure 13.28 Level transmitter.

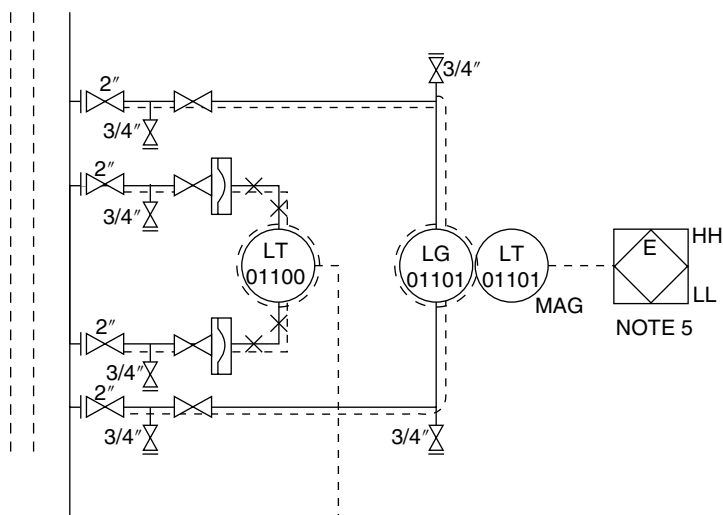


Figure 13.29 Level gauge together with a level transmitter.

There are some flow meters that have very tight clearance. It is a good practice to install a strainer upstream of such a flowmeter to guarantee the long term functioning the flow meter. One famous example of such a flow meter is a turbine type flow meter (Figure 13.31).

Tight clearances exist more commonly in high accuracy flow meters.

If a flow meter needs frequent off-line inspection, and losing the flow meter is affordable, a bypass can be provided for the flow meter (Figure 13.32).

13.11.1.5 Process Analyzers

(On-line) process analyzers are instruments that measure parameters related to the “quality” of a fluid.

There at least three types of parameters related to quality, they are:

- 1) Concentrations of individual substances like the concentration of chlorine in water, pH (concentration of hydrogen ions) of water or wet gas streams, concentration of H₂S in gas stream, etc.

Table 13.19 Full flow conductor flow sensors.



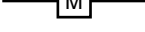

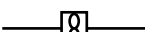



Type	P&ID schematic	Unique advantage	Unique disadvantage	Application
Orifice type		<ul style="list-style-type: none"> No moving parts 	<ul style="list-style-type: none"> Generates high pressure drop 	Most common flow measuring device
Venturi type		<ul style="list-style-type: none"> Good for low flow ranges 	<ul style="list-style-type: none"> Limited viscosity range 	Fuel measurement
Magnetic type (magmeter)		<ul style="list-style-type: none"> No moving parts Unobstructed bore 	<ul style="list-style-type: none"> Relatively high cost Fluid must be electrically conductive 	<ul style="list-style-type: none"> Water
Coriolis type		<ul style="list-style-type: none"> Can measure mass flowrate and density Can measure multi-phase flow 	Limited viscosity range	Custody transfer
Turbine type		<ul style="list-style-type: none"> Good for measuring clean, steady, low viscosity fluids 	<ul style="list-style-type: none"> The fluid should be very clean Moving parts in contact with fluid Needs very streamlined flow (long upstream straight pipe) 	Can be used for billing purposes
Vortex type		<ul style="list-style-type: none"> No moving parts 	<ul style="list-style-type: none"> Needs very streamlined flow (long upstream straight pipe) Sensitive to external vibration 	High pressure and temperature applications
Ultrasonic		<ul style="list-style-type: none"> No moving parts Unobstructed bore Bi-directional 	<ul style="list-style-type: none"> Limited solid content 	Large line sizes
Rotameter			<ul style="list-style-type: none"> The service should be fairly clean Some of them should be installed vertically 	Low flow rates
PD Meters		High turndown ratio Good for high viscosity	Mechanically intensive	Custody transfer

Table 13.20 Partial-flow conductor flow sensors.

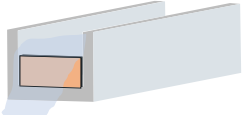
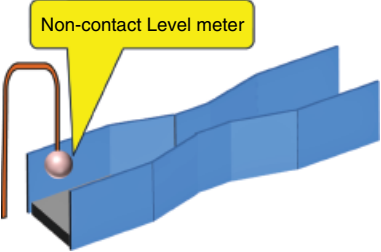
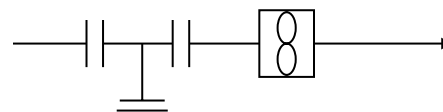
Type	P&ID schematic	Unique advantage	Unique disadvantage	Application
Weir		No moving parts		For liquid in open channels
Flume		No moving parts		For liquid in open channels


Figure 13.30 Flow sensor with high velocity requirement.

Figure 13.31 Flow sensor with very clean fluid requirement.

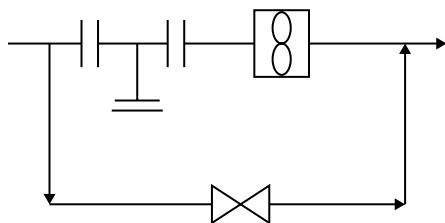


Figure 13.32 Flow sensor with bypass.

If the sensor is to measure an ion (like iron content) the sensors are named “ion selective” sensors.

- 2) Concentration of a bag of substances like TDS (total dissolved solids) in water, BOD (biochemical oxygen demand) in water, moisture content of powder solid, etc.
- 3) Indices of composition, which means parameters other than concentration but representing concentration of a fluid like reflective index, density, octane number (for gasoline streams), Brix (for sweet watery streams), etc.

The backbone of all the measurement methods is photometric methods and potentiometric methods.

In photometric methods a beam of light (or IR or UV) is emitted to the sample while in potentiometric methods electricity is sent through the sample. If neither of these two methods works, the process analyzer manufacturer may need to use more creative methods to detect the substance or parameter of interest. This is generally the cases for the third type of parameter mentioned above.

Process analyzers are generally installed on pipes (rather than tanks or vessels).

Process analyzers are not usually welcomed in process plants because of different issues. The main issues of process analyzers are:

- High cost: the justification for buying a process analyzer is difficult. A process can potentially be replaced by “operator sampling and lab test” if the parameter fluctuation is not quick and/or the respective process analyzer is not available or very expensive. Sampling systems are discussed in Chapter 18.
- They may have the cost of reagents to be filled in them periodically.
- Low reliability, although the new generations of them have less of this weakness.
- They may need frequent troublesome cleaning. However newer types have built-in self-cleaning mechanisms.

Process analyzers are built in two different types, the flow-through type and non-flow-through type (Figure 13.33). The aim of analyzer manufacturing companies is to develop process analyzers of flow-through

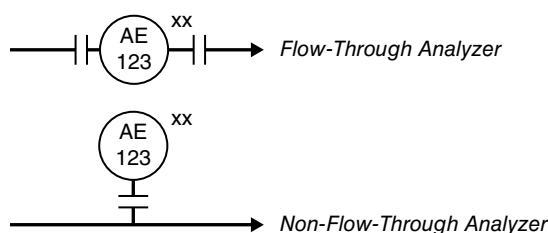


Figure 13.33 Two types of process analyzer.

type, which are a better type. However, in some cases it cannot be done and they are obliged to use the other option of non-flow-through process analyzers. The parameters that need more complicated measurement methods may need to be built in non-flow-through type process analyzers.

If a non-flow-through process analyzer should be used, an automatic sampling system is needed to direct and condition the sample to the process analyzer.

Automatic sampling systems (or samplers) are discussed in Chapter 18.

13.11.2 Transmitters

Transmitters convert weak and noisy electrical signals to good-quality transferable electrical signals. Transmitters generally need to be in the field and as close as possible to the primary instrument, so they generally don't have a divider inside their balloon. However, there are cases that a transmitter is installed in a field cabinet (double line divider in the balloon) or even in the control room (single line in the balloon).

An intermediary instrument could be a transmitter block whose function is to convert the measured signal to a suitable signal for control. The transmitter is designated “T” in the tag, e.g., “TT,” “PT,” etc. Figure 13.34 shows a transmitter block for temperature measurement.

The transmitter input signal is an electronic signal in the majority of cases and sometimes (for example in

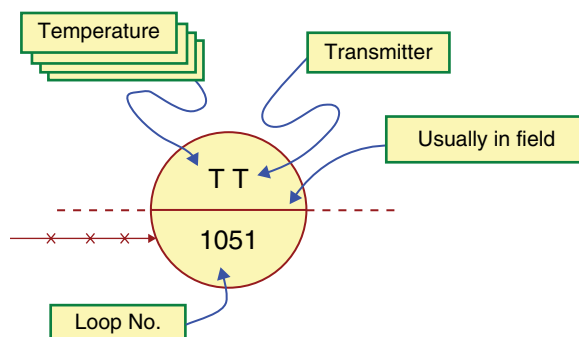


Figure 13.34 Transmitter block.

contact-type level sensors) process fluid through a capillary. There may be two incoming signals into transmitters and if not a simple single parameter is being transmitted.

The signal to the transmitter is not only from the sensor or primary element. For example in the case of flow orifice as the flow sensor, two signals go to the transmitter from pressure points upstream and downstream of the flow orifice.

The transmitter output signal is almost always electronic. This signal mostly goes to a controller but it can go to an indicator or other devices too.

13.11.3 Controllers

The function of the controller block is to compare the input signal against an SP and then generate an output signal that is proportional to the deviation. Controllers are usually located in the control room, so the tag has a divider; because they are “brains,” they are represented by a circle within a square (Figure 13.35).

The input and output signals from the controller are always electronic. They are delicate systems and should be located indoors, inside field cabinets/housings, or in the main control room.

The SP is not usually shown on P&IDs. In addition the values of the SPs are unknown and if someone is interested they need to refer to the SP table to see the value for that particular controller.

However, if the SP is a “remote set point” and comes from another control system, then it is shown as a software signal.

13.11.4 Indicators

Indicators are instruments that show parameters anywhere, even in remote area like in control rooms.

Indicators can be shown in three scenarios (Figure 13.36):

- 1) Directly on the process stream in the field (Figure 13.36(a)). These indicators are installed in the field

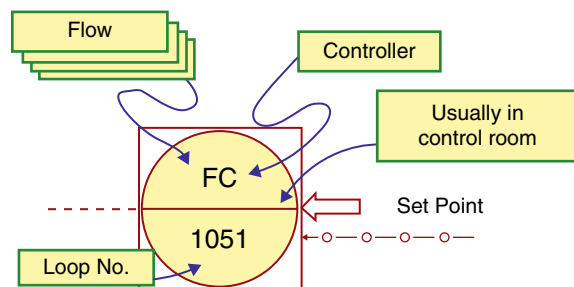


Figure 13.35 Controller block.

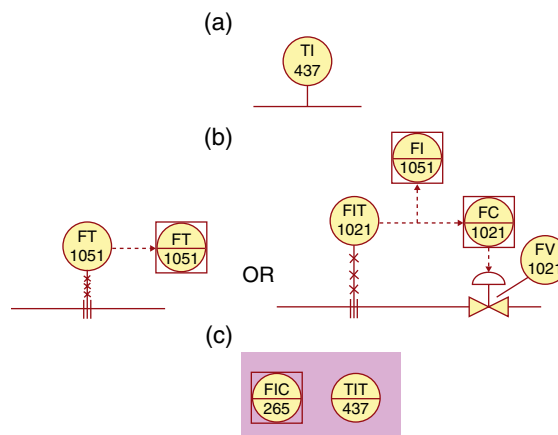


Figure 13.36 Indicators.

and are used for checking a process parameter by the rounding operator in the field. These indicators take the signal from a transmitter, even if transmitters are not shown. The tag doesn't have a divider because it's in the field. These indicators can potentially be replaced with gauges. There are plenty of field transmitters that are provided with the capability of indicating. This means the majority of XTs are XITs. There are cases that prevent us from providing an indicator in the field. One of them could be the harshness of environment.

- 2) Independently, but from the control loop and in control room (Figure 13.36(b)). These are the indicators that we use these days mainly for process parameters that need to be visible in the control room. Here the indicator tag has a divider because it is located in the control room.
- 3) As part of a block with other main functions in the control room (Figure 13.36(c)). These indicators were very popular in the past. In the early days we tried to say, “control this parameter and also show it to me in the control room.” However these days, with the implementation of HMIs (human machine interfaces), almost everything is already visible in the control room and through the monitors, even if we don't ask for it. So we no longer need to use tags like TIC or FIC.

13.11.5 Final Control Elements in a BPCS

The action of a BPCS can be either regulatory or discrete via “control loops.”

The final control element could be a variety of items but two of the most common final elements are control valves and VSDs on electric motors.

The BPCS may also handle discrete actions. This may be for a batch operation such as filtration, where we have

a filtration cycle, backwash, and then filtration again. In a batch operation, you also have direct action, for instance the switching on and off of a valve. This is not done for safety purposes but rather for process purposes. This is termed a “process interlock.” The final element can be a switching valve or the start-up or shutdown of an electric motor. These final elements in discrete actions of a BPCS are discussed in Chapter 16.

13.11.5.1 Control Valves

Control valves are basically “remotely operated throttling valves.” They comprise a throttling valve plus a modulating actuator. The valve plug can be of many types and the naming of valves is generally based on their plug type, including a globe valve, segmented ball valve, etc. There are also different types of actuators. Different types of actuators and their applications are discussed in Chapter 7.

The majority of control valves use air-driven actuators, which means they need a pneumatic signal for throttling (partially opening or closing), while all controllers generate an electrical signal. How can they communicate with each other? The same way that two people do when they want to communicate with each other and neither of them knows the other’s language; they use an interpreter. Here we need some sort of interpreter to convert the electric signal generated by the controller to a pneumatic signal that is understandable by the control valve. This interpreter is called a “transducer”, which is a signal.

For the example shown in Figure 13.35, this is shown as “LY.” “L” stands for level and “Y” means a functional modifier, an unknown character that needs clarification. In this case, it converts the electronic signal to a pneumatic pressure signal. “LY” is not always shown on a P&ID; for convenience, some companies only show an electronic signal down to the control valve, and others just show a pneumatic signal.

The final control element is a control valve. Figure 13.37 illustrates how this is shown on a P&ID.

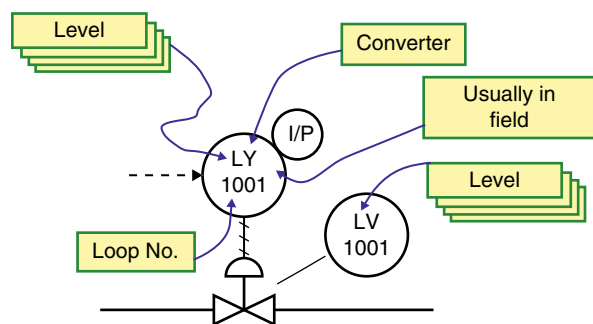


Figure 13.37 Control valve.

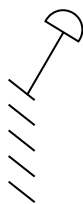


Figure 13.38 Damper.

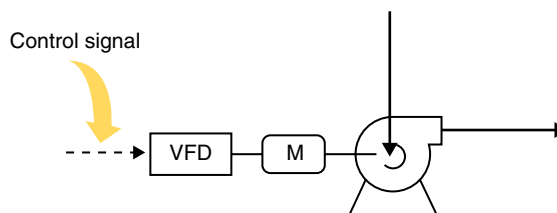


Figure 13.39 P&ID schematic of pump with rotational speed control.

However, there are some control valves that work with electrical signals (electromechanical control valves), for instance in HVAC systems. Such control valves may not need transducers.

Remotely operated dampers are a specific type of control valve for controlling the flow of low-pressure gas streams. One common application for these is the control of air flow to different pieces of equipment. The structure of a damper is very similar to that of window blinds. The actuator of the damper is connected to single or multiple blade(s) to change their angle and consequently to change the air flow rate.

Dampers (Figure 13.38) can be considered a specific type of butterfly valve that is designed for low-pressure gas streams.

13.11.5.2 Variable Speed Devices on Electric Motors

Variable speed devices (VSDs) are a group of devices that are connected to electric motors to be able to order the electric motor to decrease or increase the rotational speed of the electric motor shaft. The shaft may drive a pump, compressor or some other piece of equipment.

The best-known example of a VSD is a variable frequency drive (VFD).

Figure 13.39 shows how this method of control is shown on a P&ID.

13.12 Simple Control Loops

There are basically five different types of simple control loops: for pressure, temperature, flow, level and composition.

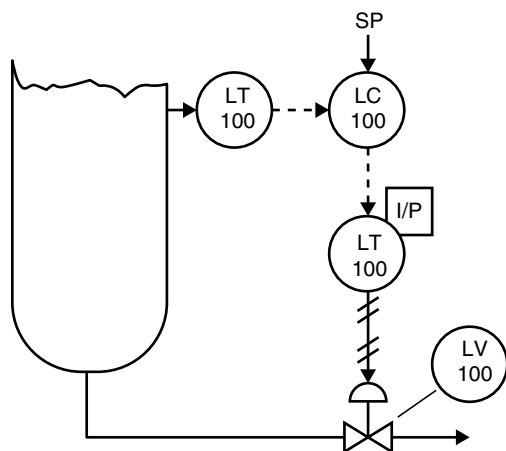


Figure 13.40 Level control loop schematic.

In the following subsection, we will go through of the concepts of each of these control loops.

13.12.1 Level Control Loops

A level control loop can be set up for non-flooded liquid containers. This means that they are applicable for tanks or non-flooded liquid vessels.

Figure 13.40 shows a schematic for a typical level loop.

As was mentioned, if the level element (sensor) is connected to the side body of the vessel, we don't show "LE" on the schematic. Instead, the schematic shows a signal going to the LT and then to the LC. Here I have indicated the SP input, but this not always shown on the P&ID. After that, the signal goes to the converter and then finally to the control valve.

Generally speaking, we always control the liquid level in non-flooded containers for the purpose of inventory control. If a container is flooded with a liquid, inventory control can be achieved by a pressure loop.

13.12.2 Pressure Control Loops

Pressure control loops can be used on pipes or on containers. Figure 13.41 shows a schematic for a typical pressure loop.

Process control loops can be used in containers and pipes and for liquids and gases. Table 13.21 shows these applications.

Pressure control loops are applicable for gases in pipes and in containers. You can think of gas pressure as similar to liquid level in tanks. Pressure control loops on gas pipes somehow shows "flow" of the pipe!

Pressure control loops are also used for liquid-flooded containers.

The use of pressure control loops on pipes is not very common; however, there are cases where we can obtain benefit from them on liquid-containing pipes.

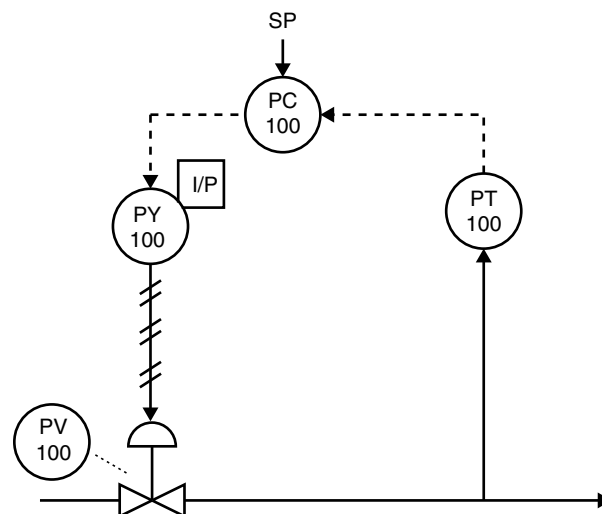


Figure 13.41 Pressure loop schematic.

Table 13.21 Application of pressure control loops.

	Liquid	Gas/vapor
Container	Only if the container is flooded	P-loop
Pipe	Not common	P-loop (or F-loop if it is around a gas mover)

Below are a few examples of using pressure control loops for liquid-containing pipes:

- To protect the downstream equipment, e.g. by opening a relieving line.
- To ensure the liquid remains in a liquid state in upstream equipment. This is important when the liquid is at a high temperature, is volatile or entering the upstream equipment at high velocity. For example, you may want to pump a liquid at high temperature using a centrifugal pump. In order to limit the damage to the pump due to gas in the line, you need to use a pressure loop upstream of the centrifugal pump to ensure the liquid doesn't vaporize.
- On utility headers. For example, on a utility water header, you may need to install a control loop to ensure that the pressure is high enough to feed the plant.

13.12.3 Temperature Control Loops

There are instances when temperature control is vital to the operation of a particular piece of equipment. Examples are furnaces, boilers, heat exchangers and temperature-fixed reactors.

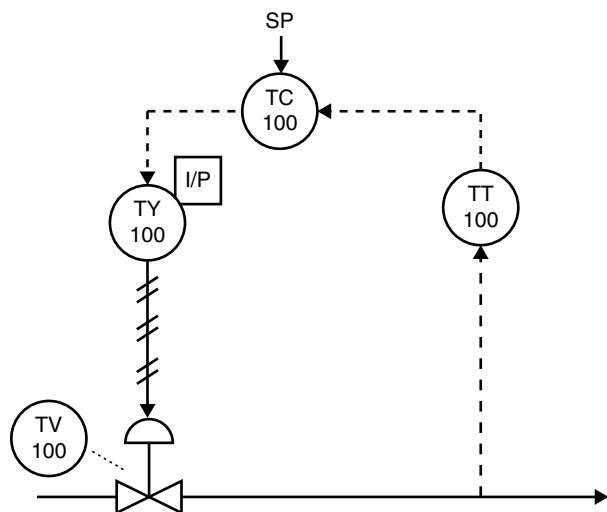


Figure 13.42 Temperature control loop schematic.

Figure 13.42 shows a typical temperature control loop, with a signal going from the sensor to the transmitter, then to the controller, then through a converter and finally to the control valve. Temperature loops are generally placed on “energy-carrying streams.” There are two main energy-carrying streams: heat transfer medium streams and the fuel streams.

Heat transfer medium streams can be steam, cooling water, or hot or cold glycol, and are generally provided by utility systems. Fuel streams include natural gas that is used as fuel for burners in fired heaters and boilers.

Temperature control loops can be very sluggish. For example, when controlling the temperature in a room via an HVAC system, the lag time between the temperature sensed and the room temperature reaching the SP of the controller can be long. Since the reaction times of temperature loops are not very fast, they are prime candidates for cascade control (discussed in Chapter 14).

Sometimes when we want to control composition we need to use a process analyzer. Process analyzers are very expensive and, because of their unreliability, they are not often incorporated in plant control. We may use temperature control instead of composition control in many cases because they are often interrelated. This type of control is called “inferred control.”

For example, you may want to separate a mixture of alcohol and water using a distillation tower. Because alcohol is more volatile than water, the alcohol will be drawn off the top of the vessel and the water off the bottom. To check the purity of the alcohol, we could install a composition control loop, which would adjust the operating temperature of the distillation tower accordingly. In this case, there is a direct relationship between the operating temperature of the vessel and the purity of the

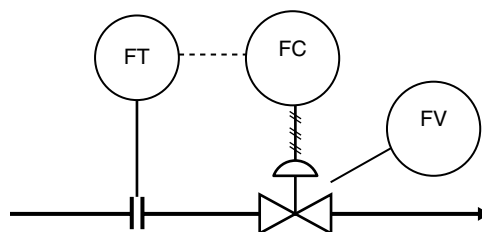


Figure 13.43 Flow loop schematic.

alcohol. So, in effect, we can substitute a more reliable temperature control loop for the composition loop.

13.12.4 Composition Control Loops

Composition loops are used in unit operations that are used to alter the composition of the process stream, e.g. a reactor, distillation tower or fractionation column. However, as I’ve mentioned before, we generally try to avoid process analyzers in control loop applications unless the process analyzer is adequately reliable. It is more common to see process analyzers in control if the process analyzer is of flow-through type.

Composition control is even more sluggish than temperature control. Once again, when employing composition loops we may use a cascade control structure.

13.12.5 Flow Control Loops

Figure 13.43 shows a typical control loop schematic for flow, which we only use on pipes. The signal from the flow element on the pipe goes to the transmitter, then to the controller and finally to the control valve.

Flow control loops are only used on pipes and not on any other process items. They may be used for liquid flow or gas flow in pipes.

13.13 Position of Sensor Regarding Control Valves

When facing a control loop on a pipe, the following question may arise: “where should the control valve be located, upstream or downstream of the sensor?”

Figure 13.44 shows possible configurations of control loops.

The answer to the above question depends on the process parameter we are sensing.

For example for a flow sensor, the answer is: “we always put the control valve downstream of the flow sensor in flow control loops.” This is because the control valve will generate some disturbance in the flow and the flow element will not be able to measure the flow rate accurately.

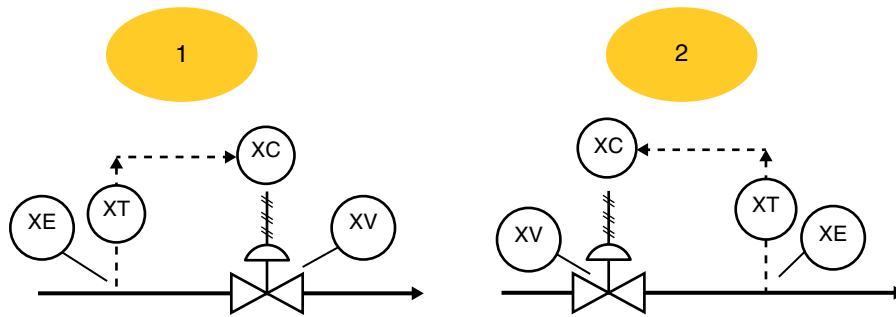


Figure 13.44 Control loop configurations.

You can eliminate the effect of the disturbance in flow from the control valve by locating the flow element far downstream, but this will make the control loop very slow. Good design practice dictates that the flow element should be upstream of, and as close as possible to, the control valve.

For other process parameters the answers are as below:

- **Pressure.** For pressure control, both configurations 1 and 2 are acceptable. It depends on whether you want to control the process based on the upstream or the downstream pressure. If the pressure sensor is upstream of the control valve, the name of the loop is

the (fore) pressure loop. If the pressure sensor is downstream of control valve, the name of loop is the back pressure loop.

- **Temperature and composition.** For both of these parameters, either configuration can be used. Generally speaking, any disturbance caused by the control valve does not influence the performance of temperature or composition sensors.
- **Level.** For level, the question can be changed to whether the container should be upstream or downstream of the control valve. The answer depends on the decision for plant-wide control. This concept is discussed in Chapter 15.

14

Application of Control Architectures

14.1 Introduction

This chapter is not meant to be a definitive course in control system design. Instead, I want to take you through the process principles involved in designing a system. There are two main methods that can be used:

- Design by analysis. This method is used when the unit operation is complex enough that it requires mathematical equations and chemical process data to render a solution. For instance, this method would be used by control engineers in the design of the control system for a fractionation tower or distillation column. In these operations, where composition is a vital process parameter, you may have many side streams that draw off different end products at various stages in the vessel. So you can imagine that this requires quite a lot of analysis to design the control system correctly.
- Design by intuition. This method involves a mixture of gut feeling, practical experience, and observation to provide a control solution. This is the way I learned, and it is also the method used by most designers. This method can be used because most process operations are not complex enough to warrant a full mathematical analysis for control purposes. Apart from that, many items of equipment like pumps, heat exchangers or boilers have established a control methodology that works and has been tried and tested over many years, so it can be learned easily.

In this chapter, we focus on the second method: designing a control system by intuition.

14.2 Control System Design

There are four steps involved in designing a control system:

- 1) Selecting the parameter you want to control and the location of the sensor.
- 2) Identifying the manipulated stream, or the stream on which you want to place a control valve.

- 3) Determining the set point.
- 4) Building the control loop.

14.3 Selecting the Parameter to Control

In this step, you are basically selecting the type of sensor and its location by identifying the process variable you need to control. In the majority of cases, the sensor should be placed on the stream whose process parameter you want to control.

Later you will see the two main different types of control loop architectures, which are “feedback” and “feedforward.” The above statement is valid only for feedback (FB) loops. It means in feedforward (FF) loops, the sensor does not necessarily need to be located on the stream that is to be controlled.

How do we go about selecting the right parameter to control?

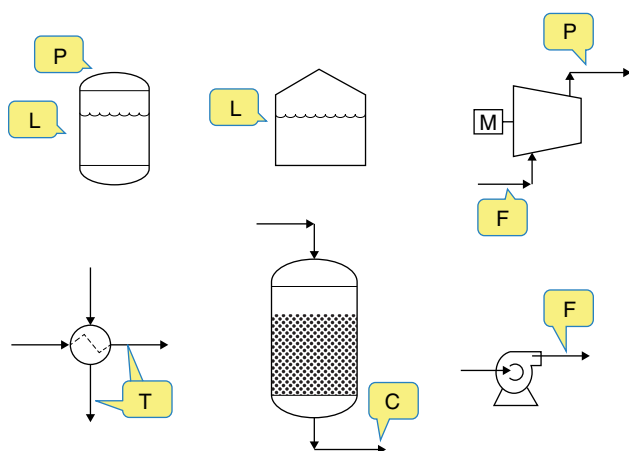
The approach that I take is to use Table 14.1 as a rule of thumb to help with parameter selection.

This table is useful because, for each parameter, it gives you examples of where the sensor should be placed and also the point in the process where the parameter must be adjusted. There are a few points to note about this table:

- 1) Always control the inventory in your process.
- 2) Pressure for gas vessels works similar to level for liquids. Both of them work for inventory control.
- 3) Do temperature control wherever there is a piece of equipment that causes a change in temperature.
- 4) Do composition control wherever there is a piece of equipment that causes a change in composition.
- 5) Composition control is not common for at least two reasons: one, because their sensors (process analyzers) are slow and not very reliable, and two, because composition is generally a function of other parameters and by controlling temperature, pressure, etc. the control of composition can be achieved.
- 6) Flow rate is only controlled in piping systems (obviously), and control loops are located near fluid movers.

Table 14.1 Sensor location.

Parameter	Sensor on:	Examples
Level	<ul style="list-style-type: none"> Wherever there is a liquid in a non-flooded container 	Tanks, non-flooded vessels
Pressure	<ul style="list-style-type: none"> On gas containers On gas movers On pipes (gas or liquid, but mainly for gas) to adjust pressure for pressure-sensitive equipment 	Gas vessels, gas space of a separator, compressors, fans, blowers, gas stream to an HVAC heater, hydrocyclone
Flow rate	<ul style="list-style-type: none"> On pipes and generally around fluid movers 	Pumps, compressors, fans, blowers
Temperature	<ul style="list-style-type: none"> Wherever there is a change in temperature 	HX, furnaces, boilers
Composition	<ul style="list-style-type: none"> Wherever there is a change in composition 	Reactors, unit operations (evaporators, etc.)

**Figure 14.1** Examples of sensor location.

- 7) It is not rare to see a container as part of/inside a unit operation or process unit. In such cases, controlling the unit operation and unit process involves inventory control too.
- 8) Always control the pressure for unit operations and process units whose function depends on pressure. Examples are flash drums, two-phase flow separators, and hydrocyclones.

Figure 14.1 shows some examples of sensor locations:

- In a non-flooded pressurized vessel, both pressure and level may be important parameters and you may choose to control both of them.
- In a storage tank for a liquid we only control level. Pressure variation in liquid storage tanks is very small.
- For a heat exchanger, obviously temperature is the controlled parameter. However, only one temperature is important: the outlet temperature of the stream of interest (or target stream).
- In a reactor packed with a catalyst, you need to control the operation by analyzing the composition of the process stream on the outlet line.

- For fluid movers, we can choose to control either flow or pressure because they are inter-related. For pumps, we always control flow. For compressors, we can either control the flow or the pressure. If we decide on flow, we locate the flow element on the suction side for economic reasons because it is cheaper than buying a low pressure design flow sensor (for the low-pressure side of the compressor). If we control pressure, we can place the element either on the suction pipe or on the discharge pipe because the cost difference is not significant.

14.4 Identifying the Manipulated Stream

This step involves the identification of the stream where we want to install our control valve, called the manipulated stream.

Control engineers use the term “pairing”; they want to find a good pairing of the controlled parameter with the manipulated stream. This procedure involves selecting the stream for which a change has the most direct, the strongest, and quickest effect on the parameter you want to control (the target parameter), with minimum interference with the other parameters.

When identifying the stream to manipulate, there are four main factors to consider with regard to how the selected manipulated stream will influence the controlled variable:

- 1) Directly. Will the manipulated stream influence the target parameter directly? For example, if you put a control valve on the inlet pipe to a tank it will influence the level inside the tank directly. An important rule of thumb to follow if you want to avoid a sluggish control system is to locate the control valve on the manipulated stream as close as possible to the sensor for the controlled variable.
- 2) Strongly. Put your control valve on the stream that will influence the target parameter most strongly.

For example, you may have more than one flow coming into a tank. The control valve should be installed on the line with the highest flow rate because that will have the biggest influence on the target parameter (the level in the tank).

- 3) Rapidly. Which manipulated stream will produce the fastest response and influence the target parameter most rapidly? Sometimes you are confronted with a dilemma here; the response of one stream may be fast but its influence on the target parameter may not be as strong as another stream with a slower response. The rule of thumb here is: the stream with the stronger influence wins over the one with a quicker response.
- 4) With minimum interference with other parameters. Whenever you install a control valve, it can cause a disturbance to other parameters and adjacent equipment. The control valve needs to be located where it will cause the least disturbance to other parameters and units.

The fourth factor is always extremely important to take into account; don't leave this step out! In some instances when the control loop is in operation, the disturbance caused by the control valve may cause other process parameters to fluctuate. In such cases, the system cannot be controlled by conventional strategies. A more advanced system, called a "model predictive control" (MPC) system may be necessary. The design of an MPC is not covered in this course.

One important rule of thumb regarding placing control is as follows:

"Don't put more than one control valve on liquid-filled pipelines. A vessel that is flooded with liquid is considered as a piece pipe with liquid in it (from a control standpoint). There are a few exceptions to this rule, but these cases are very rare. On gas pipes there may be more than one control valve, but they cannot be very close to each other."

14.5 Determining the Set Point

There are two types of set points that are used in process control: a manual set point (MSP) and a remote set point (RSP).

Manual set points for most process variables can be obtained from:

- The heat and mass balance (H&MB) data for the process.
- Manufacturer's or equipment design data.

In theory, we can extract all of these set points from the documents and use them to control the plant.

However, in practice this would make the control system too "rigid" and extremely difficult to operate.

What we do is to introduce remote set points generated by the plant, which give the control system more flexibility and allow the system to "float." An RSP is generated by another control loop, and this process of communication between different control loops is critical for the smooth operation of the plant.

Just as it is not ideal to control a plant with MSPs exclusively, it is also not good to use only RSPs. This would lead to the process going completely out of control. You have to have a certain number of MSPs in the control system in order to conform to the overall H&MB of the plant. Non-conformance would lead to an inefficient, uneconomical process that would manufacture a product of inferior quality.

Let us look at our options for the application of MSPs and RSPs:

- Flow. Here we can use both manual and remote set points. The value for the MSP is obtained from the H&MB. MSPs for flow loops are mainly for loops at the "edge" of the plant (at the beginning or at the end), and not in the middle of units.
- Temperature. Set points for temperature control can be manual or remote, depending on the application. There is a rule that is specific to heat exchangers. Heat exchangers can be classified as process heat exchangers if both streams in the heat transfer operation are process fluids, or as utility heat exchangers if one of the streams is a utility such as steam or cooling water. The rule is: with process heat exchangers, it's better to go with an MSP obtained from the H&MB and with utility heat exchangers, you would normally go with RSPs.
- Pressure. In this case, you can use either manual or remote set points. The value for the MSP is available from the H&MB.
- Level. For level, we only use an MSP, which is the normal liquid level (NLL) of the container. We generally don't use a remote set point to control the level in a container.
- Composition. In the case of composition, we always use an MSP. This value is obtained from the H&MB.

The above concepts are summarized in Table 14.2.

As a general rule, we try to maximize the number of remote set points and minimize the MSPs for optimal flexibility in process control. However, we cannot do away with MSPs completely because they are necessary to ensure that the process conforms to the overall H&MB of the plant.

Table 14.2 Set point application.

	MSP	RSP
Flow	Available from H&MB table	Available from other loops
Temperature	Available from H&MB table	Not very common practice
Pressure	Available from H&MB table	Available from other loops
Level	Available from NLL of container	Not very common practice
Composition	Available from H&MB table	Not available (not meaningful)

14.6 Building a Control Loop

In this section, I want to discuss different structures and architectures for control loops:

- Feedback versus FF control, or a combination of the two.
- Multiple-loop control.

14.6.1 Feedback Versus Feedforward

There are two main types of continuous process control concepts: feedback control and FF control. The majority of control loops in industry are of FB type.

What is the difference between FB and FF? It is basically the difference between being reactive or proactive.

To illustrate this, we can look at two groups of people: one group who eats lunch when they're hungry, and the other, who eats at 12 o'clock, whether they're hungry or not. The group that eats when they're hungry is reactive; they are reacting to a signal, or FB, from their bodies that they are hungry. The other group, which eats at 12 o'clock, are being proactive; they are saying, "let us eat at 12 o'clock to prevent ourselves from getting hungry later on," and they could be called FF people.

The features of each group are listed in Table 14.3.

These principles can be applied to process control; FB loops take corrective action after an upset is propagated, while FF loops take corrective action before an upset is propagated.

Table 14.3 Two ways of handling hunger.

Group 1	Group 2
People who eat lunch when they are hungry	People who eat lunch because it is 12 o'clock
Feedback people (FB)	Feedforward people (FF)
Reactive	Proactive

Table 14.4 Feedback versus feedforward control.

Feedback control	Feedforward control
Taking corrective action after upset propagated	Taking corrective action before upset propagated. It "predicts" the disturbance and proactively takes action to prevent it.
Adjusting an error after occurrence	Preventing error before occurrence
More popular	Less popular, only for important or critical loops usually in combined form However, one specific type, ratio control, is very common wherever there is a chemical injection system.
Default control system	<ul style="list-style-type: none"> • When change in controlled value can make a big, non-affordable disturbance • Can be useful if all the disturbances are known and quantifiable. In other words: it should be designed by a very good designer! • "Blind" control; cannot check the result of the action
Closed loop control	Open loop control

But what is best, FB or FF? Theoretically, FF control should be better than FB control because it is proactive. However, in practice the overwhelming majority of control loops operate on FB. Why is this?

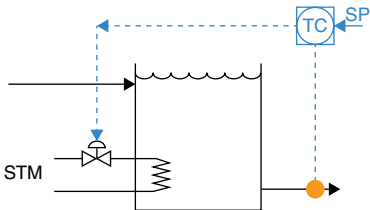
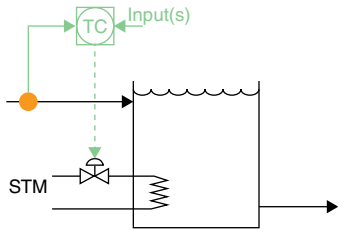
The reason is that if you want to be a "good," proactive person you need to understand all the events in order to prepare yourself for them. In the process control world we say, "feedforward control is perfect when it is designed by a perfect engineer!" As there is no perfect person in this world who is able to foresee all of the possible fluctuations in the control loop design, we deviate from the concept of FF.

Based this analogy, we can define FB and FF controls as shown in Table 14.4.

Nonetheless, we still use FF loops for some cases. For example, in very simple loops where the nature of fluctuation(s) are well known, or where a fluctuation needs to be taken care of at the beginning before its full propagation happens. The other example of application of FF loops is in loops that are related to the economy of a plant, such as the yield of a process. For the cases where we deal with "difficult" processes we may need to use FF control. "Difficult" control in this context means processes that are very swift (like pH change, which is logarithmic) or processes with wide range of variation, or processes that the needs-to-be sensed parameter is difficult to be sensed accurately and quickly.

Now is the time to bring a technical example. Let us look at a practical example to see the difference between

Table 14.5 Schematic of feedback versus feedforward control.

Feedback control	Feedforward control
 <p>Sensor on: resultant stream (generally downstream of equipment)</p> <p>Control valve on: stream affects the parameter that has sensor on it</p> <p>The majority of control loops in process plants</p>	 <p>Sensor on: fluctuating stream (generally upstream of equipment)</p> <p>Control valve on: stream affects the parameter that has sensor on it</p> <p>Not very common in industry and very rare as standalone control for a piece of equipment (if needed they are used in the combination form with feedback control)</p>

the function of an FB control loop versus FF type. See the example below (Table 14.5) about warming up water in a tank using a coil of steam.

Table 14.4 shows the two different modes of temperature control of a liquid in a tank.

The temperature of the liquid in the tank is held by a steam heating coil.

The FB control system measures the temperature on the discharge line, compares it to the set point and then adjusts a control valve to regulate the flow on the steam line to the vessel. Around 90% of control loops in industry operate on FB.

In FF control, the sensor is located on the feedline to the tank. The control engineer must use a mathematical formula to include all process variables, predict and control the outlet temperature based on the fluctuating input and then adjust the control valve on the steam line accordingly. This is a true proactive approach, which leaves no room for error, and for this reason most people prefer to have the backup of FB control as well. So, you very rarely come across a pure FF control system. The combination of FF with FB control is far more popular.

In the example, you can see the difference between FB and FF in the P&ID document:

- There is no difference in the location of control valves.
- If you want to control temperature, the sensor should be a temperature sensor in an FB control loop, but this is not the case in an FF control loop. The parameter being controlled is not always visible from an FF loop.
- In an FB loop, the sensor is located downstream of the fluctuation while in an FF loop, the sensor is located upstream of the fluctuation. Hint: this is different from

what some people mistakenly say: “in an FB loop, the sensor is located downstream of the equipment, while in an FF loop the sensor is located upstream of the equipment.” even though this interpretation could be true in plenty of cases.

When reading a P&ID, the way to differentiate between an FF and an FB loop is to see where the sensor is located. If it is located on the fluctuating stream then it is an FF system. An FB loop will have the sensor located on the resultant stream.

14.6.2 Single- versus Multiple-Loop Control

Single-loop control is the default mode for system control. In other words, whenever you can, keep it simple and don't overcomplicate the control system. As we have discussed, single-loop control may be in the form of FB or FF control.

Multiple-loop control can have a number of different architectures:

- Cascade control
- FB + FF control
- Ratio control
- Selective control
- Override control
- Split range and parallel control

It is worth mentioning that each control loop in a multi-loop control architecture can be separated from the other control loop(s) and work as a single independent control loop, if this is desired during the operation of a plant. Such capability is generally for the control room operators.

14.7 Multi-Loop Control Architectures

Up to now we have learned the architecture of a single loop control. In this section we will focus on the architectures of different multi-loop control systems.

14.7.1 Cascade Control

There may be some cases where we have to deviate from a simple, single control loop and use cascade control.

Figure 14.2 shows a simple loop for level control on a tank, and how we could improve this by moving to cascade control.

In the left-hand schematic, there is a simple level loop on a container. Because of some “reasons,” we need to move away from a simple control loop to a complicated cascade loop. In a cascade loop, the level controller output (and not the level sensor output) is used as the set point for a flow control loop on the pipe.

To be able to talk about different features of cascade control we need to learn some terminologies of cascade control. Let’s look at (Figure 14.3).

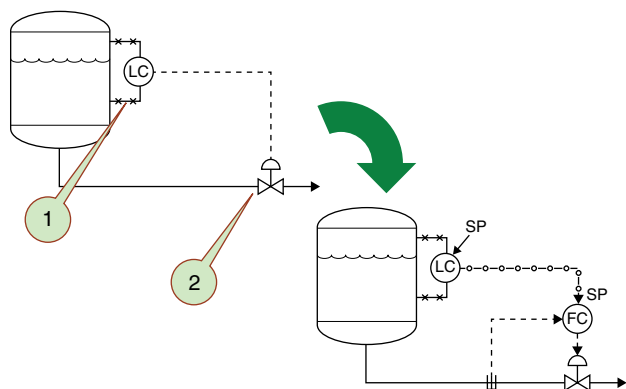


Figure 14.2 Moving from a simple loop to cascade control.

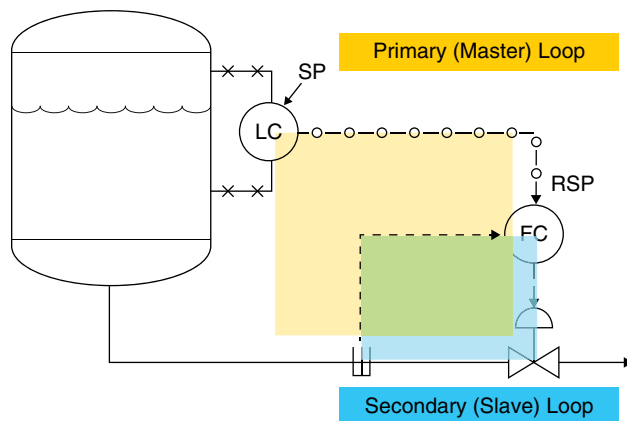


Figure 14.3 Cascade control terminology.

Another name for cascade control is reset control. The main loop is called the primary, or master, loop. The second loop is called the secondary, or slave, loop. We refer to the control architecture for this tank as level-to-flow cascade control.

The “reasons” that forced us to move away from a simple control loop can be lumped into two groups: some of them located in the container (group 1 problems) and some of them located on the controlled pipe (group 2 problems).

These reasons are listed here. First four of these reasons lie in the area of the sensor and one of them (the last one, number 5) lies in the area of the control valve.

- 1) Large fluctuations. The controlled variable in the primary loop may have large fluctuations, making it difficult to control.
- 2) Tight control. You may need to keep tight control over a narrow band of operation for the controlled parameter. One example could be control in a small vessel. Irrespective of the process, the level control of a small vessel needs to be tight. If tight control is needed, we may be forced to use cascade control.
- 3) Slow response. The response to control action on the parameter may be slow. A parameter could be inherently slow, like temperature, or it could be slow in a specific application. For example, temperature sensors in large spaces are very slow, or, when a sensor is very far from the control valve the loop would be slow.

Figure 14.4 shows the relative speed of parameters.

If we are dealing with slow loops, we may have to use cascade control rather than single-loop control. In such an arrangement, the slow parameter should always be in the master loop, and the faster parameter will be in the slave loop.

- 4) Response rate. In some cases, we may want to limit the rate at which the controlled variable increases or decreases.
- 5) Severe fluctuations in flow in the secondary loop. There could be a case where the control valve is placed on a pipe with severe flow fluctuations. In such cases, using cascade control is inevitable. One famous reason that a pipe may experience severe fluctuations is when that pipe is a header or

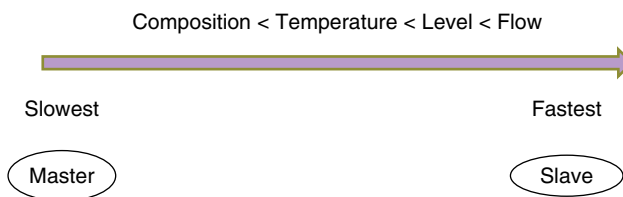


Figure 14.4 Sluggish control system.

manifold feeding multiple users. The other causes of potential high surge flow are:

- Existence of a container with limited capacity (more likely a vessel) at the beginning or at the end of a pipe (especially if the containers are controlled by level control loop).
- Very long pipe.

By implementing cascade control, we can exercise better control over the process variable.

In cascade control architecture, the primary loops are generally composition and temperature, which have a bad reputation for sluggishness. The secondary loop could be flow or pressure, but in most cases it is flow because it has the fastest response of all the process parameters.

Figure 14.5 shows several types of cascade controls that are common.

Is it true that by using cascade control we can control two parameters at once? Meaning, when you have level-to-flow cascade control, can we say we that both level and flow are controlled at the same time?

The answer is no!

Each control valve can control only one parameter. If there is an effect on a second parameter because of operation of a control valve, this is generally called “interference” and does not have a predictable effect.

The above answer is from a purely theoretical standpoint. However, in practice, if the variation of the master parameter is somehow limited, we can say that “both of parameters are fairly well controlled.”

In the example of level-to-flow cascade control, if the cascade control is installed on a fairly small vessel, the variation in level is very limited and we can say: “although the main intention of level-to-flow cascade control is to control the level tightly, because level variation is already low, the set point for the flow loop has a small variation and therefore we can say the flow is almost free of variations.”

Let’s have a look at some examples of cascade control.

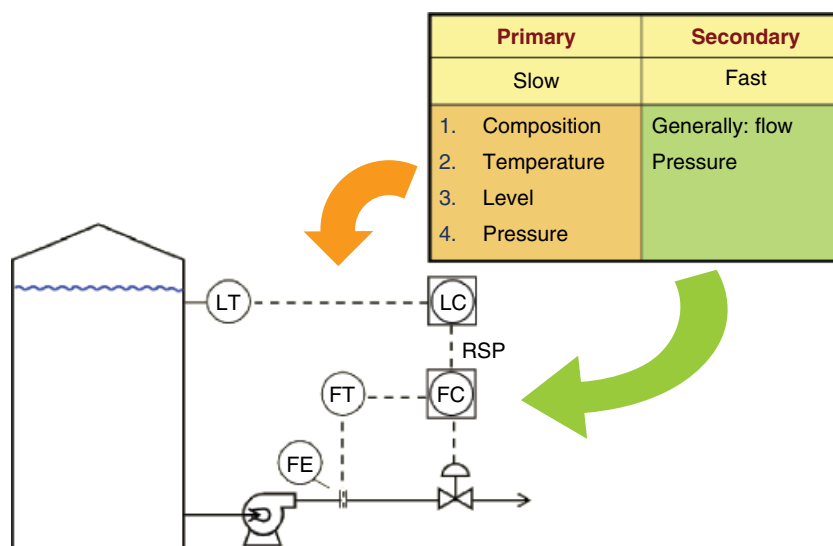
Figure 14.6 shows a distillation column, for separating alcohol from water. At the top, we want to draw off pure alcohol. In the first schematic we have a process analyzer, AE, going to a transmitter, AT, then to a controller, AC and finally to a control valve. As we know, composition has the slowest response of all parameters so it is better to go with cascade architecture to control the purity of the alcohol.

The next schematic, Figure 14.7 shows the composition generating a set point to a secondary loop, which controls the feed to the distillation tower. This cascade control helps to fasten the loop.

The final schematic, Figure 14.8, shows the introduction of a temperature control loop as well. So here we have a three-tier cascade system, which you don’t see very often. Why introduce a temperature loop? Well, as you may remember, we said that composition is measured by a process analyzer, which is not only slow but also unreliable. For this reason, a temperature loop is introduced as backup for the composition loop for more effective control of the distillation column.

Let’s consider another example of cascade control.

Figure 14.5 Common cascade control architectures.



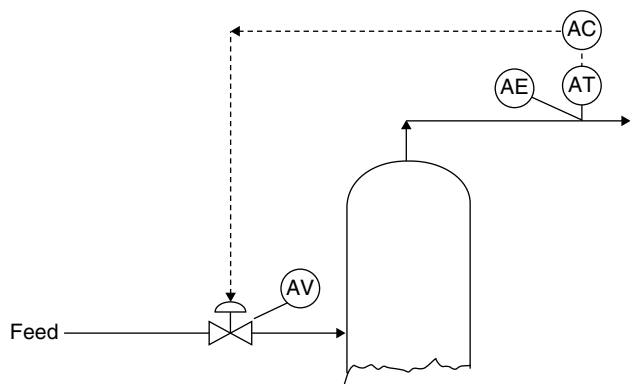


Figure 14.6 Simple composition control architecture.

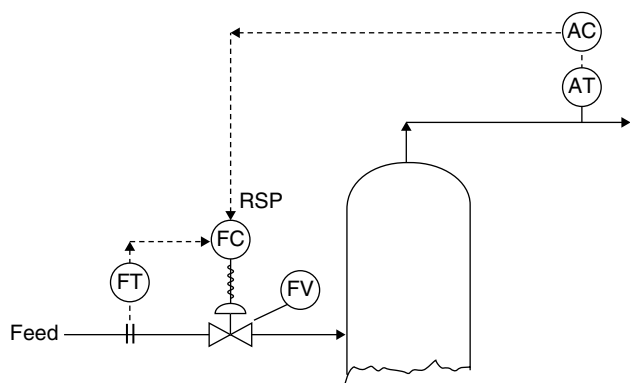


Figure 14.7 Composition-to-flow cascade control architecture.

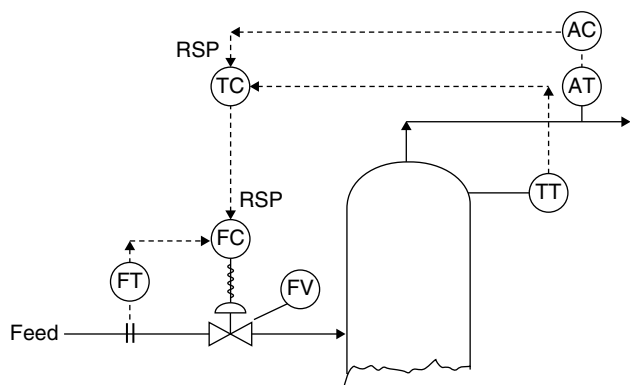


Figure 14.8 Composition-to-temperature-to-flow cascade control architecture.

Figure 14.9 shows a conventional drum boiler that has two drums; the top drum is called a steam drum and the bottom one is called a mud drum. For this operation to work, we need to control the level in the steam drum because if the level drops too low the drum may be burned and be destroyed. If the level is too high, we will have carry-over of liquid into the steam line. The combination of a small drum and the need for tight control

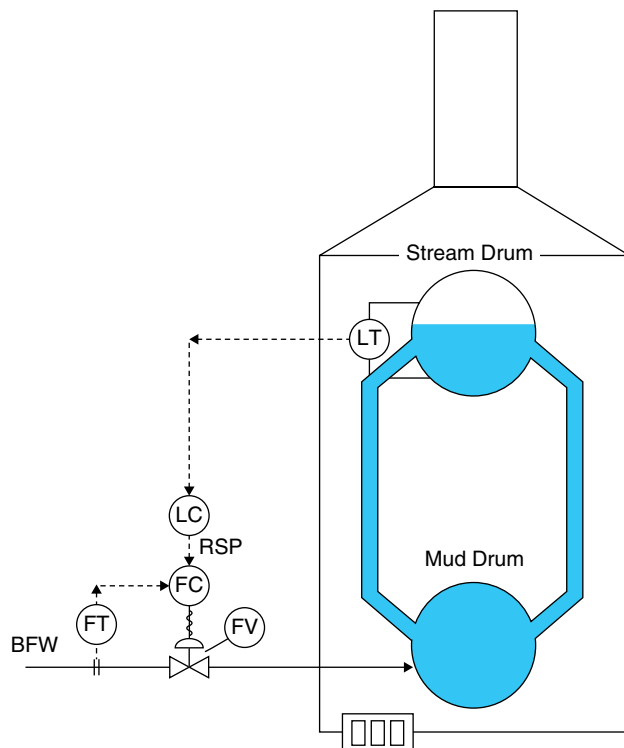


Figure 14.9 Level-to-flow cascade control in a steam drum.

means that we need to install cascade control. We call this level-to-flow cascade control.

Before finishing cascade control, let me share with you a practical point:

Cascade control is basically a “non-luxury” version of combined FF–FB control!

When we are looking for a superior and tight control, we may think of FF control (in addition to FB control). However, implementing FF control is not easy and we may decide to use cascade control instead.

14.8 Feedforward Plus Feedback Control

As was mentioned, an FF control loop is not a good control practice in the real world. Actually, in the real world only a few (or none!) purely FF control loops may exist in each plant. If there is an FF control loop, there is a high chance that there will be a “combined” version of that: FF plus FB.

There are at least two ways that you can design a combined FB with FF control system, as shown in Figure 14.10.

First, in the schematic Figure 14.10(a) the FB signal can be used to adjust the FF signal to the control valve. This arrangement is called “feedforward compensation,” or “feedback trim.” Some P&IDs will show this as a Sigma sign in a circle with + and – on either side of it, but you don’t always see it shown that way.

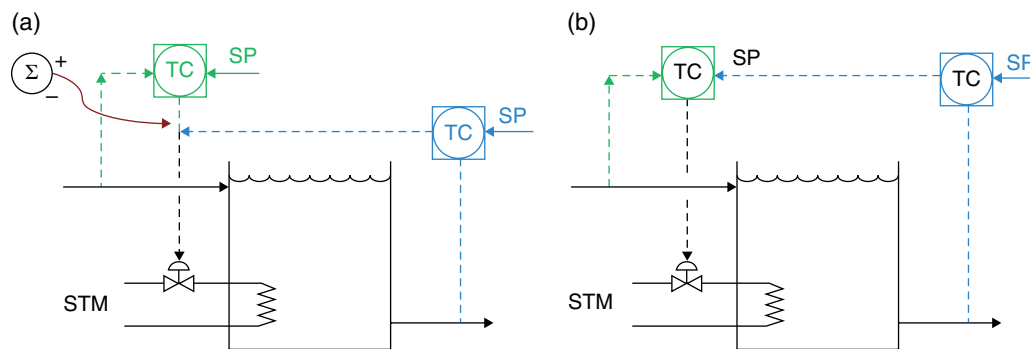
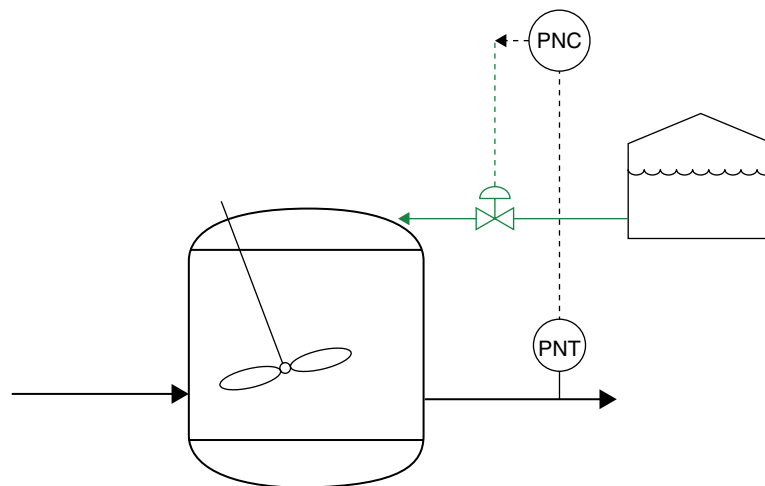


Figure 14.10 Different types of FF + FB control.

Figure 14.11 First attempt to control a neutralization vessel-feedback control.



Second, in the schematic Figure 14.10(b) the FB signal can be used as a set point for the controller on the FF loop. In this arrangement, the FB loop determines the set point for the FF loop. This arrangement is very similar to cascade control loops.

One thing to remember is that the FF loop is always the main “driver” of the system in any type of FF + FB combination.

A very good example of an FF + FB control system is a GPS in a car when you are fairly familiar with the route to get to your destination. You set the GPS with the address of your destination, and off you go. This is the FF loop, with the GPS as the controller. If you make a wrong turn along the way, the FB loop will inform you (and the GPS), which will then use your new position as a set point and recalculate your route to get back on track to your destination.

Let’s look at an example in a neutralization tank.

The following schematics show various control mechanisms for the neutralization of water in a tank. The first schematic, Figure 14.11, shows a classic FB control loop, with the sensor located on the resultant stream and a signal to a control valve on the line from an acid/base tank to control the desired pH in the water tank.

The second schematic, Figure 14.12, shows an FF control loop for the same operation. There is no difference in the position of the control valve – it is still located on the pipe from the acid/base tank. However, the sensor element is situated on the inlet stream to the water tank, and sends a signal to the controller. The controller will calculate an adjustment based on a mathematical formula, $f(x)$, in this case a titration curve generated in the laboratory. It will then send a signal to the control valve.

Since we can’t always rely on the accuracy of the FF system, it is better to use a combination of FF and FB control. This is shown in the final schematic in Figure 14.13.

In some cases, you will find FF + FB in combination with cascade control, as seen in Figure 14.13.

This concept is not very difficult to understand: “if a parameter is so important that it deserves FF control, then it deserves cascade control too!”

The schematic below shows the bottom of a distillation tower, where we need to control temperature. The temperature controller adjusts a control valve on the steam pipe. However, because temperature control is very slow,

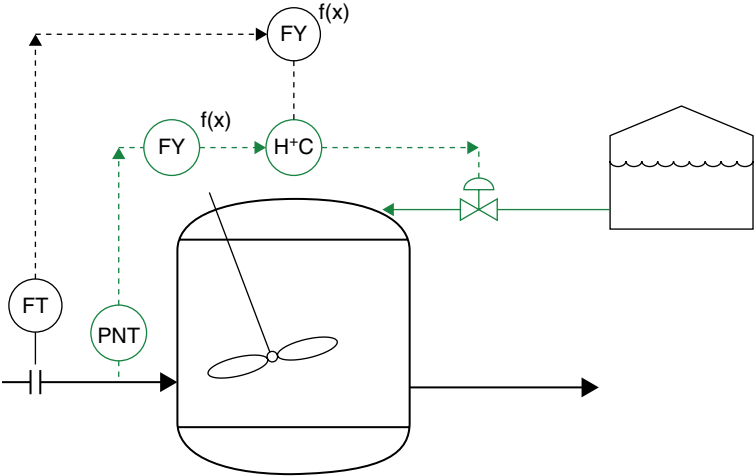


Figure 14.12 Second attempt to control a neutralization vessel-feedforward control.

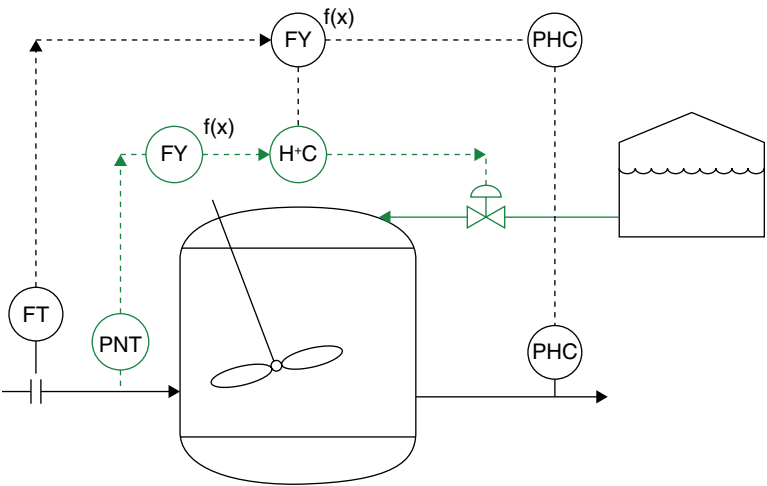


Figure 14.13 FF + FB control for neutralization vessel feedback plus feedforward control.

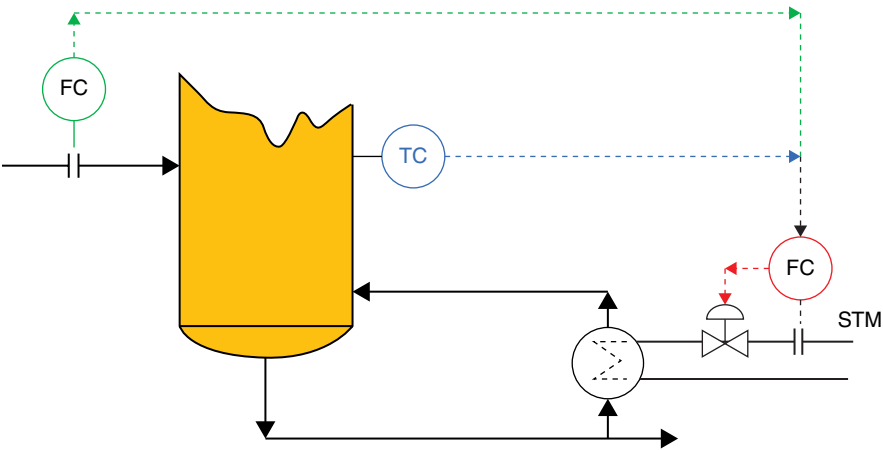


Figure 14.14 FF + cascade + FB control.

we add in a temperature-to-flow cascade system. We also have an FF loop (the top flow loop), which provides a set point for the cascade system.

14.8.1 Ratio or Relationship Control

In ratio control, the set point of a loop is determined by another sensor signal. Be careful: it is determined by another sensor's signal, not by another control loop (or controller signal). The stream with the control loop on it is called the “controlled stream,” and the stream that provides the set point is called the “wild stream.”

Although a ratio control could be theoretically defined for any pair of parameters (even non-same parameters like flow pressure, or flow temperature), they are most common for flow–flow situations.

Flow–flow ratio controls are used very often for mixing, blending and chemical injection systems.

The wild stream may require the controlled stream to have the same flow as itself, or a fraction of its flow, a multiplier of it, or some non-linear mathematical function of it. As in the majority of practical cases, a fraction or multiplier is used in this type of control. This is called “ratio control,” but it may use a complex function and, because of this, a more suitable name for this type of control could be “relation control.”

Ratio control is inherently a type of FF control. This highlights the importance (or necessity) of adding an FB component if the control could be impacted by unknown disturbances.

There are two well-known schemes that are used in ratio control.

In “type-one” ratio control, the wild stream flow sensor parameter is “processed” in an “operator” to get the required flow rate in the controlled stream, which is then used as the (remote) set point in the control loop of the controlled stream. In this type of ratio control, the “process variable” signal is the flow signal of the controlled stream.

In “type-two” ratio control, the flow sensor signals from two streams are fed into an “operator” and “processed” to get one “compounded” flow sensor signal, which is used as the sensor signal (and not the set point) in the control loop of the controlled stream. In this type of ratio control, the set point is an MSP and is determined based on a mathematical relationship between two streams.

The most popular ratio control is type one. A schematic of this is shown in Figure 14.15.

This schematic shows two pipes. The sensor on the top stream (wild stream) sends a signal to the flow modifier, which uses a function, $f(x)$, to modify the signal and this then becomes the set point for the control loop on the

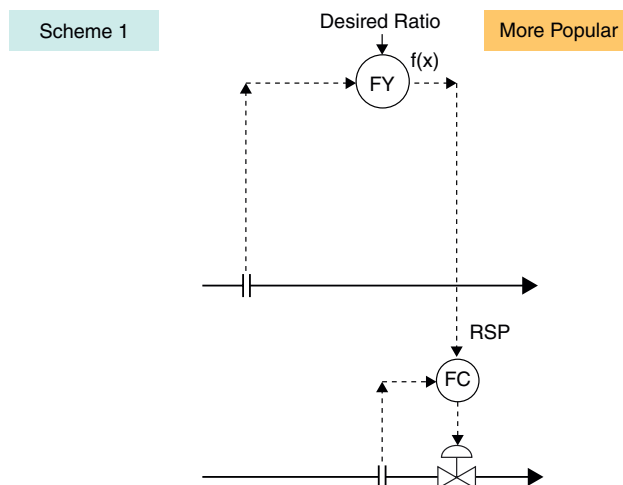


Figure 14.15 Role of ratio control.

other stream. As we mentioned, $f(x)$ can be a ratio or some other mathematical function.

In P&IDs, this type of ratio control can be shown differently from that in Figure 14.15 by using FFC elements instead of an FC element. An “FFC” element, or “flow fraction” element, is a symbol that shows the combined version of FY and FC (flow signal operator and flow controller). This type, showing type-one ratio control, is shown on the right-hand side of Figure 14.16.

It should be noted that what is seen in Figure 14.16, on both the left-hand and right-hand sides, are only two ways of “showing” this concept, and there is no difference in the physical arrangement of these two representations.

Figure 14.16 shows how we can merge the flow modifier, FY, and the flow controller, FC, and label it as FFC on the P&ID. Again, this stands for “fraction flow controller.”

In “type-two” ratio control, we have signals from sensors on both streams going to a flow modifier. This uses a dividing function to divide the one value by the other. It then sends the modified signal as a set point to the flow controller. In this instance, the flow controller acts like an FFC because it receives a set point signal that is a fraction. Figure 14.17 shows one example of “type-two” ratio control in practice.

You may recall that I mentioned that ratio control is inherently an FF system, and that, when we discussed FF and FB control, an FF loop works much better in combination with an FB loop. This is shown in Figure 14.18.

For example, this could be a blending system with ratio control on the two streams. We then install an FB loop from a composition sensor on the blended line through the composition controller, AC, which in turn sends a set point to the flow controller, FC. This is a perfect example of ratio control combined with FB.

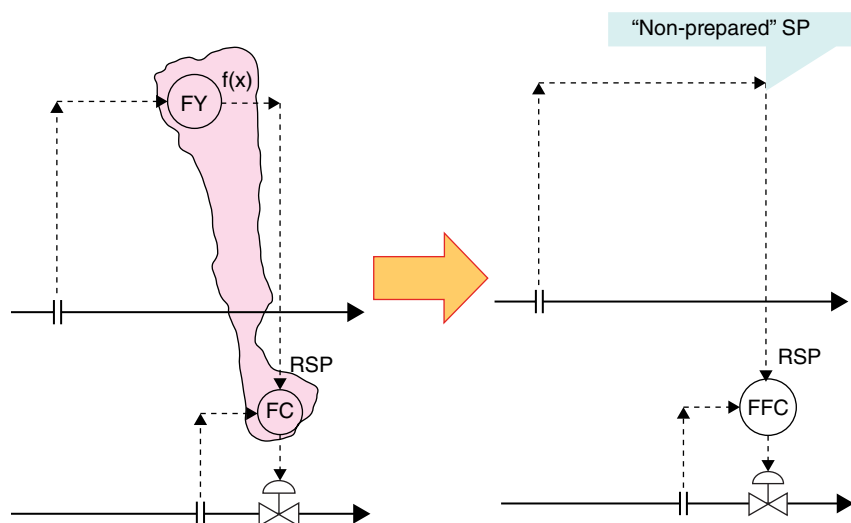


Figure 14.16 Ratio control symbology.

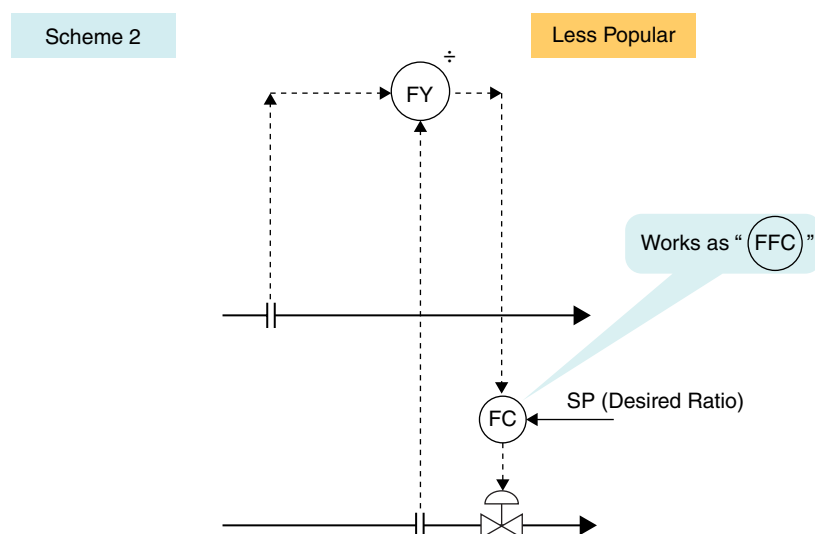


Figure 14.17 Ratio control: alternative scheme.

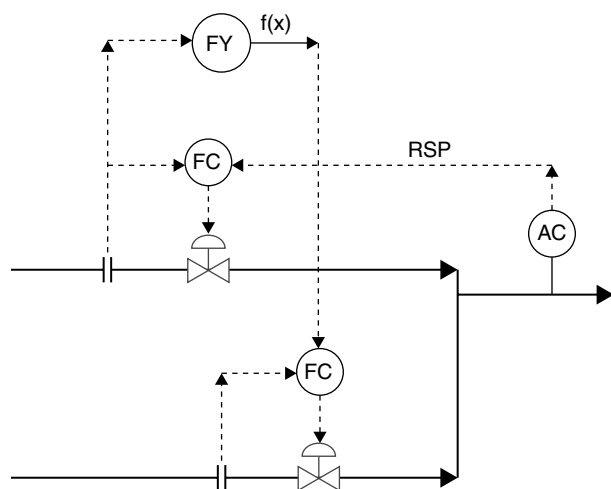


Figure 14.18 Ratio control: FF + FB.

14.8.2 Selective Control

When we talk about selective control, we mean that we want to select one signal from multiple signals to be processed by the controller. We can do this by using either a high selecting (> or HS) or low selecting (< or LS) operator, which is situated on a primary signal (sensor signal).

Selective control is used where there are several sensors installed to measure one unique parameter of a single stream, or “comparable streams,” and one single sensor signal needs to be “selected” to be sent to the controller for control purposes.

The selective control could select the highest value, lowest value, or a mid-range value as the “selected sensor signal” for the controller.

“Comparable streams” are streams that are completely identical. Generally speaking, there are no “comparable

streams” in a process plant, except parallel streams to identical parallel process units. You may ask: “why would one install more than one sensor for one specific parameter?” The short answer is unpredictability. If, for whatever reason, you are not sure where on a stream you can find your “representative” value of the parameter of interest, you may use selective control.

Let’s look at an example of a reactor, shown in Figure 14.19.

The feed stream comes into the reactor from the top and the product leaves from the bottom. The operation involves an exothermic reaction and the reactor is jacketed so that we can cool it down. The coolant enters the reactor at the bottom and leaves at the top.

Because we are unable to predict where in the reactor the temperature will be very high, we place a number of temperature elements at various places on the reactor with temperature transmitters that send signals to a high selector, TY. The selector then selects the highest temperature (worst case), and sends a signal to the temperature controller, which in turn adjusts a control valve on the coolant feed line.

Figure 14.20 shows another example of selective control on two operating centrifugal pumps. Here we need to protect the centrifugal pumps against low flows, i.e. flows that are less than the “minimum flow” reported by the pump manufacturer. With centrifugal pumps, if the flow drops below the minimum specified by the manufacturer, the pump will start to vibrate.

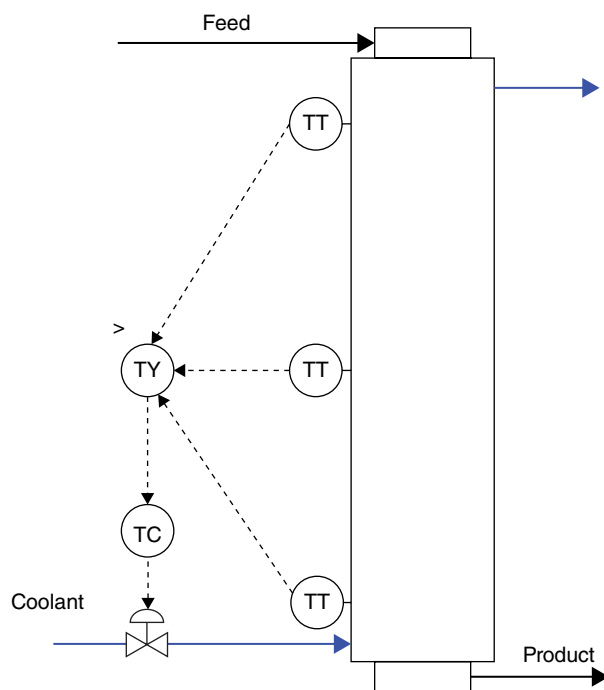


Figure 14.19 Selective control.

In this case, we have two pumps operating in parallel (This is not a case of one pump being on standby; they both operate at the same time). In order to protect the pumps from minimum flow, sensors from both pump discharge lines send signals to a low selector, FY. This in turn sends a signal to the flow controller, FC, which adjusts a control valve on the recirculation line to ensure that the flow into the pumps stays above the minimum.

14.8.3 Override and Limit Control

Override and limit are two different types of control; however, both of them use high and/or low functions to operate. Therefore, it is a good idea to compare them side-by-side before we look at each one in more depth.

First of all, we need to discuss the difference between selective control (discussed previously) and override control. Override control uses the same operators as selective control: a high selecting (>) or a low selecting (<) operator. However, the main difference between them is that selective control is part of the normal operation of the control system acting on primary signals from sensors, whereas override and limit control only kick in on controller signals when the process drifts outside of its normal band of operation.

At the schematic level, in selective control the operators “sit” on top of the sensor signal while in override control, the operators sit on top of the controller output.

Now it is time to talk about the difference between override control and limit control. Table 14.6 highlights the differences between override and limit control.

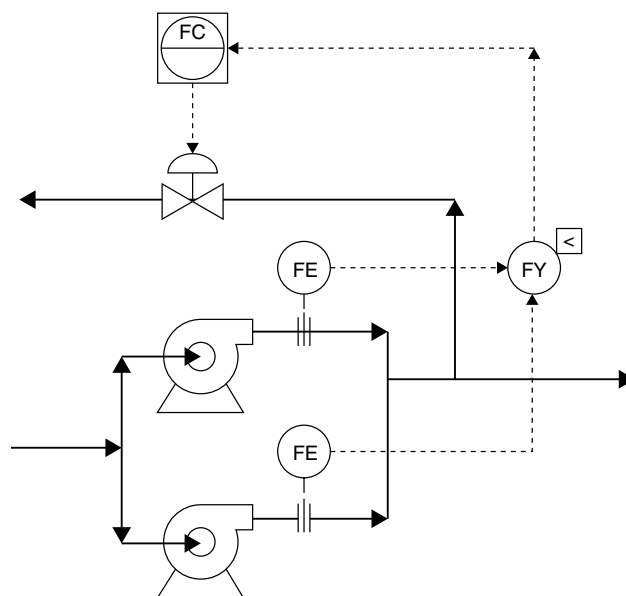
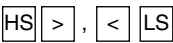



Figure 14.20 Selective control example.

For example, we can liken this to flying an airplane. If the main pilot is lost due to a heart attack, you have two options: the co-pilot takes over or, if he is also out of operation, you go to autopilot. The co-pilot is the better option and can be likened to override control in a process, where at a certain point the “pilot” will be switched. Going to autopilot is like limited control in a process, where at a certain point you go into protection mode.

In override control, we have signals from two control loops taking care of one control valve. However, the control valve is selective in that it only sees one of these signals, depending on the situation.

Table 14.6 Override versus selective control.

	Override control	Limit control
Symbol		
Name	High/low selecting	High/low limiting
Number of in-signals	Two	One
Existence of built-in input	No	Exists
Structure	Here is two control loops on one control valve, but which loops work on the control valve at any time depends on the situation	Here is one control loop. After a certain point, there is no longer any control (signal is limited)
Airplane example	After certain point, “pilot” will be changed	After certain point, the airplane goes to “protection” mode

Limit control operates on one signal, limited to a specific value, coming from one control loop and is only activated at a certain point, when there is no longer any control.

Override control is less invasive to the process than limit control.

It should be noted that if it is decided to implement both override control and limit control in a specific system (which is not usual), it should be in such a way that the sensitive point for override control is placed before the sensitive point for limit control.

This concept is shown in Table 14.7.

Limit control is a type of control – as part of BPCS – that can be implemented to avoid “handing over” the process unit to SIS functions. The “constraint point” of limit control is a bit below the trip point for the SIS function.





Now the question that should be answered is: what is the difference between high and low in these two different types of control?

Table 14.8 shows the difference between high and low operators for override and limit control. A high-selecting operator/high-limit operator chooses the larger signal; this is used in a process that is sensitive to low parameter values. On the other hand, a low-selecting operator/low-limit operator chooses the smaller signal in a situation where it is dangerous to operate the system at high values.

It is not unusual to see more high-selecting operators or high-limit operators (> or \nless) in systems because generally systems are more sensitive toward the higher level of parameters.

Now it’s time to talk about override control and limit control separately.

Table 14.7 Action points for override and limit control.

	Override control	Limit control
Symbol		
Effect in process	Less invasive for the process	More invasive for the process
Action range	<div><div>----- Mechanical protection</div><div>----- Hi Hi</div><div>----- Hi</div><div>----- Normal operation</div><div>----- Low</div><div>----- Low Low</div><div>----- Mechanical protection</div></div>	<div><div>----- Mechanical protection</div><div>----- Hi Hi</div><div>----- Hi</div><div>----- Normal operation</div><div>----- Low</div><div>----- Low Low</div><div>----- Mechanical protection</div></div>

14.8.3.1 Override Control

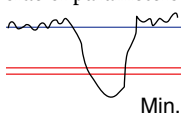
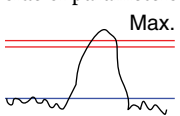
The concept of override control can be summarized as follows:

When the control loop is moving toward its borderlines, it may start to disturb another parameter. In such cases, the would-be disturbed parameter (in the form of another control loop) can be “added” to the primary loop by a selector switch to release the original control loop from the burden of controlling and bringing the new, wise control loop (overriding control loop) on to control the system.

In the above definition, there are different phrases that need more clarification:

- The borderline of a process parameter could be any level in the band from HLL to LLL, but they are generally closer to the HLL or LLL threshold.
- The “disturbance” can happen because of excessive opening or closing of the control valve in the primary control loop. “Disturbance” means upsetting the smoothness of the process. If the disturbance is something related to safety, the case must be handled by the SIS system and not by override control.
- The selector could be a high (>) or low (<) selector.

Table 14.8 High versus low operators.

Override control	\lt High-Selecting	\lt Low-Selecting
Limit control	\gt High-Limiting	\lt Low-Limiting
Duty	Picks larger signal. “Floor keeper,” sensitive to low side of parameters	Picks smaller signal. “Ceiling keeper,” sensitive to high side of parameters
		

For each override control system, two control loops can be identified: the normal acting loop and the overriding loop. The override signal is basically a wise signal that takes care of the control duty when the normal signal starts getting a little crazy.

It is not always clear on the P&ID which is the normal operating loop and which is the overriding loop. The way of recognizing this is by studying the loops’ process concepts. In some P&IDs, there is a note saying: “overriding signal” to denote it.

It is important to recognize the normal acting loop and the overriding loop during development of P&IDs.

The last important thing to consider is that the selector sits on top of the controller outputs (COs), and not on the sensor signals.

Let us look at some examples of override control.

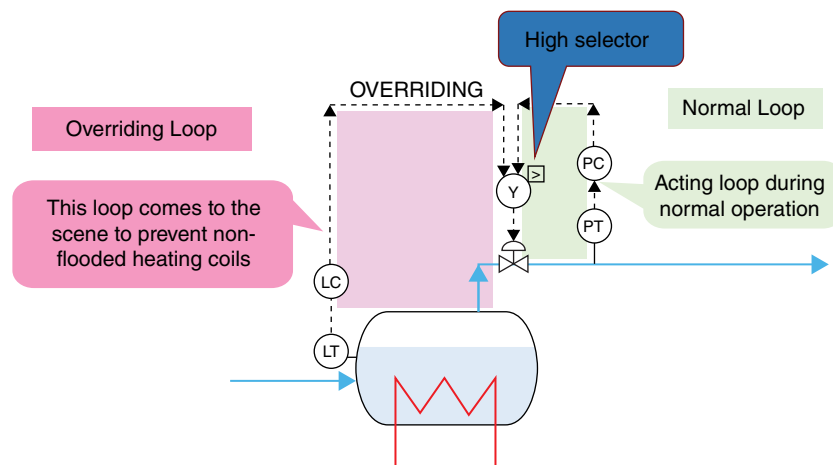
Figure 14.21 could be a steam generator. The first thing to notice is that we have an operator (<) on a controller signal, and not a sensor signal. This indicates that we have override and not selective control.

We should have two control loops: a normal acting loop and an overriding loop.

During normal operation, the acting loop, which is the pressure loop in this case, is in control. When there is an increase in demand for steam from the process downstream, the pressure drops and control valves open up to allow the vessel to generate more steam. When there is less steam demand from the downstream process, the control valve pinches back to decrease steam generation.

Now think about a case where steam demand goes excessively high. In such a case, the control valve may go to fully open and an excessive amount of steam will be generated. This could lead to a drop in water level and a consequent problem if the heating coil is left dry. So we introduce an overriding loop to control the level and prevent an exposed heating coil.

Figure 14.21 Example of override control: steam generator.



In this case, we implement another loop as a level loop, and combine it with the existing pressure loop by a selector function. We basically tell the pressure control loop: “if you are going crazy and opening your valve widely and/or over a very long time, we will bring in a wise level control loop to make sure the heating coil won’t be left naked and won’t be burned out.”

Now the question that should be answered is: which one of the selectors should be placed here – the high selector or the low selector?

There are two ways to answer this question: with a simplistic approach and with a precise approach. In the simplistic approach we can say: “Since the override control tries to limit the control system from a ‘low’ level of liquid, we need a ‘floor protector,’ or high selector (>).”

For the precise answer, we need more information about the hardware of the control system.

Let’s see what happens when the override control comes on to the scene.

Because we are not happy with that “low signal,” we pick the high selector (>). We basically picked the high selector because the signal that was “messing up” was a low signal.

As seen in Figure 14.22, we show that a low signal to controller generates a low signal to the control valve. This is true only in direct acting control valves. Here we – apparently – assumed the controller is a “direct acting controller.”

The other question is: “who said the low signal after the controller makes the control valve more open (and not more close)?” Here we – apparently – assumed the control valve is air-to-close type, but it could be air-to-open.

In essence, our previous logic is based on some assumptions, which are “direct acting controller” and “air-to-close control valve” (Figure 14.23).

These assumptions are not always true though.

Let’s see the case where we have a reverse acting controller in Figure 14.24.

Here you can see there is no problem with “high demand steam,” but we will have a problem with low steam demand, as shown in Figure 14.25.

In this case, we need to use the low selector (<), because we are not happy with a high signal.

We know whether the valve is air-to-open or air-to-close.

- A fail closed (FC) valve is an air-to-open valve.
- A fail open (FO) valve is air-to-close valve.

In the simplistic approach, we ignore all these complexities, and this may be acceptable at the P&ID level, since at that level we don’t know the details of the controller type (direct acting or reverse acting).

In another example, shown in Figure 14.26, we have a compressor controlled by a pressure loop, shown as the blue loop, which is the acting loop during normal operation.

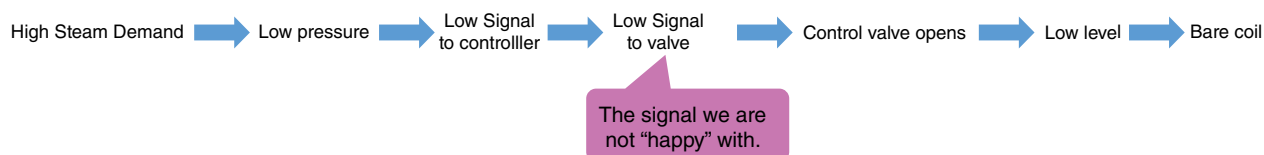


Figure 14.22 Deciding on the selector type in an overriding control system.

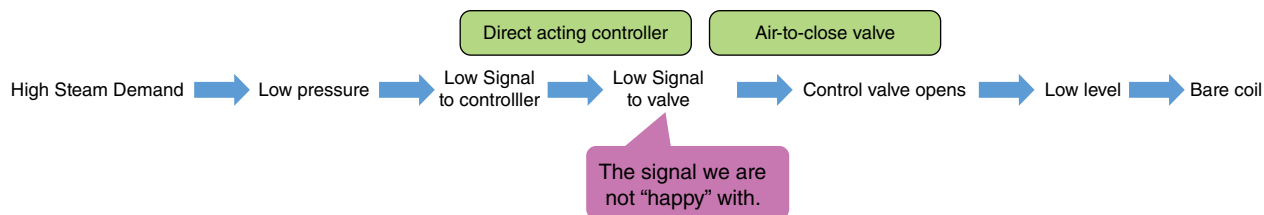


Figure 14.23 Assumptions in deciding the type of selector.

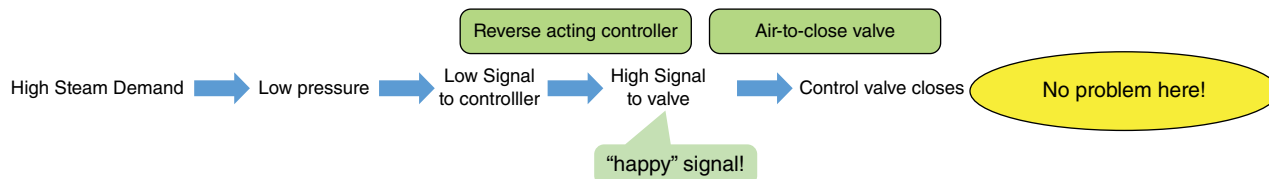


Figure 14.24 Another set of assumptions in deciding the type of selector.

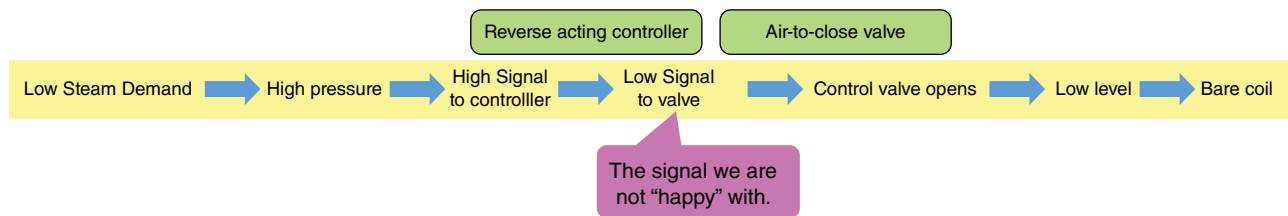


Figure 14.25 Changing the selector type based on a new set of assumptions.

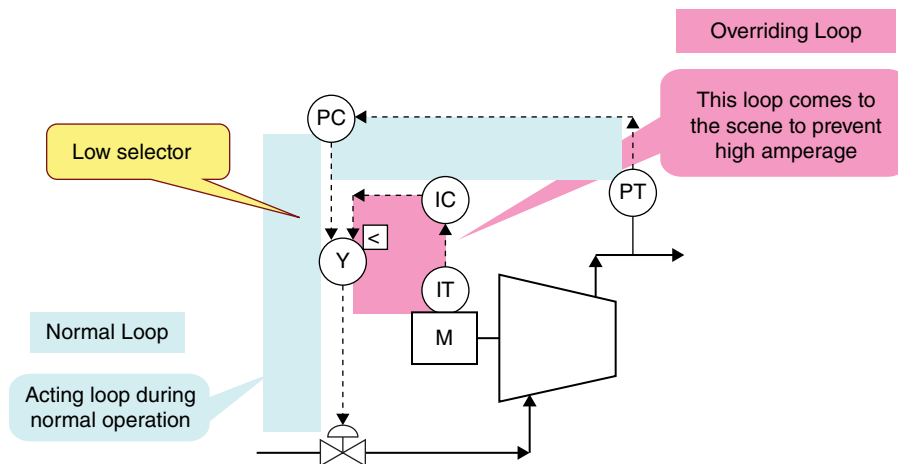
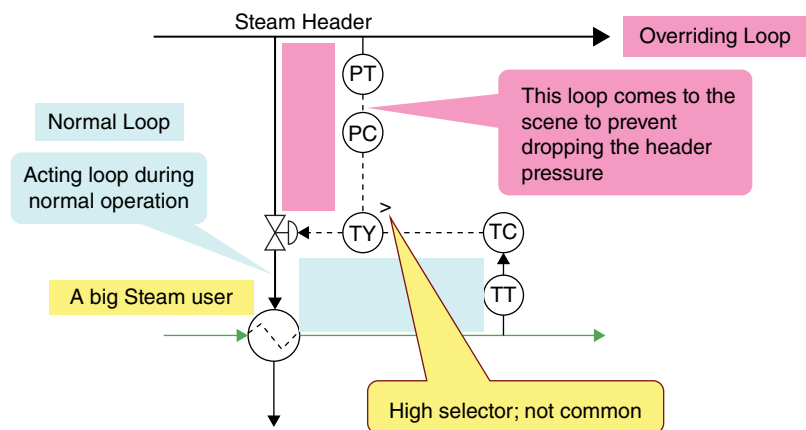


Figure 14.26 Example of override control: compressor.

Figure 14.27 Example of override control: steam header.



A compressor is a very expensive piece of equipment and the higher the flow to the compressor, the higher the load on the electric motor.

We don't want the motor to burn out, so we put in a current controller, IC, with an override on the signal from the pressure controller to the control valve in order to protect the motor. The override control will activate when the amperage on the motor goes above a set level.

In the next example (Figure 14.27), we have a steam header that provides steam for a number of users. One of them, a heat exchanger, is a much bigger user than the

others. The acting loop during normal operation is a temperature control loop that adjusts a control valve on the steam line to the heat exchanger.

Because it is the biggest user, if this heat exchanger draws too much steam from the header, the pressure in the header will drop and the other users will not have access to steam. A pressure loop is necessary to override the signal from the temperature controller to the control valve. This is done to maintain pressure in the steam header.

Another approach to override control is a way to combine two control loops if they are not important simultaneously.

You may have two different control loops, each with a control valve that competes for control. You can merge the two control valves into one and then rank each controlled variable as a basis for override control.

In this context, override control is basically an attempt to control a system when the single loop generates interference. In such cases, we generally say the solution is to implement “model predictive control” (MPC), but override can be used in very simple cases.

Override control is used primarily for optimization or automatic start-up of units. Sometimes it is implemented in units to upgrade and improve them if the units show frequent process upset, equipment failure, trip, or PSV actions.

Override control is also used for cases where the magnitude and/or extent of the disturbance cannot be predicted accurately. Some example are flows coming from underground (oil and gas extraction) or flows in utility networks with multiple users.

Actually, the second example is a very common example of the application of override control: where a utility consumer is suspected to be using more of the utility than expected and it may drop the pressure of utility stream (and impact several other utility users), an override control may be implemented to isolate the “badly behaved utility user” from the utility network and protect the other utility users as soon as its usage goes beyond a reasonable level.

The other examples of using override control are listed in the Table 14.9.

14.8.3.2 Limit Control

Limit control can be considered to be another form of override control. However, the override signal doesn't come from the plant; it is manually set at a particular value inside the controller. If that value is reached then limit control is activated.

A limit signal is a fixed and wise signal, which takes care of the control duty when the primary signal goes to its crazy side. It is not used in most plants, but its main application is when start-up and shutdown of plant is automated, or in the normal operation of highly automated plants.

Table 14.9 Process cases that call for the application of override control.

To prevent process upset	To maintain equipment integrity
<ul style="list-style-type: none"> • Drop in utility network pressure (for big users) • Fouling • Creation of two-phase flow 	<ul style="list-style-type: none"> • Drop of NPSHA and cavitation • Fire heater tube burnout

Limit control is used mainly for plant safety and optimization, but it can be replaced by an SIS.

Figure 14.28 shows an example of limit control. In this example, we have a temperature control loop for a furnace. The temperature controller adjusts a control valve on the air feedline to the furnace. A limit control value of 180°C is set on the temperature controller to make sure the internal tubes won't burn out.

14.8.4 Split Range and Parallel Control

We learned that to design a control loop we need to find one stream as a manipulated stream to put a control valve on. There could be cases where more than one flow rate affects a single “process variable” (PV). In such cases, we are normally able to select the one that affects the PV most strongly and eliminate the others from consideration. However, where you have two streams that are equally influential, you need to implement split-range or parallel control through control valves on both process variables (both streams).

Split-range or parallel control is introduced where you have more than one manipulated stream/one control valve that you need to control. This is most prevalent in pressure loops on gas streams, but it has been implemented on liquid streams as well.

In parallel control, both control valves work simultaneously together while in split control, one control valve starts its work when the other control valve starts to reach its end point (closed or open), and doesn't move when the second control valve is functioning.

For both parallel and split control, there are two modes: straight and reverse. In straight mode, both control valves move toward opening or closing (simultaneously or

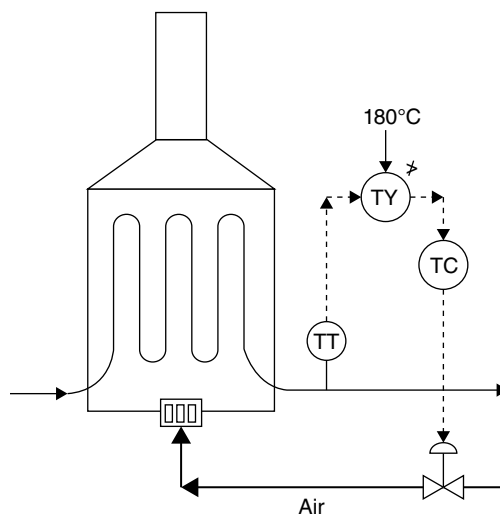


Figure 14.28 Example of limit control in a furnace.

non-simultaneously), but in reverse mode, while one control valve is going toward opening, the other one is going toward closing (simultaneously or non-simultaneously).

Table 14.10 gives a graphic illustration of the difference between split-range and parallel control operating on two control valves.

Therefore, there could be four different modes of “multi-valve” control. Here we explain two of them, and the two others are easy to interpret.

Let’s look at the split-straight control type. When the process parameter is at its normal level, both control valves are fully open. This means that CV1 (control valve 1 on stream 1) is fully open, and CV2 (control valve 2 on stream 2) is fully open too. When the process parameter starts to deviate from its normal level, CV1 starts to close, and CV2 remains open until CV1 closes fully. At this point, CV2 starts to close. CV1 remains closed while CV2 is operating.

Now let’s look at the parallel-reverse control type. In this mode of control, CV1 is fully open and CV2 is fully closed when the process parameter is at its normal level. Then, when the process parameter deviates from normal level, CV1 starts to close while at the same time CV2 starts to open.

When reading a P&ID, we need to make sure to understand which types of control are used out of the four

types we have introduced. Not all P&IDs will indicate whether the control is split-range or parallel, or whether the operating mode is straight or reverse. The most complete P&IDs show a diagram below the control system to show the intent of the control.

One distinguishing difference between split and parallel control is that for split control, the middle point (X%) must be mentioned on the P&ID.

However, the mode of control – straight or reverse – is generally not mentioned on P&IDs.

Now let’s see some examples of parallel/split control.

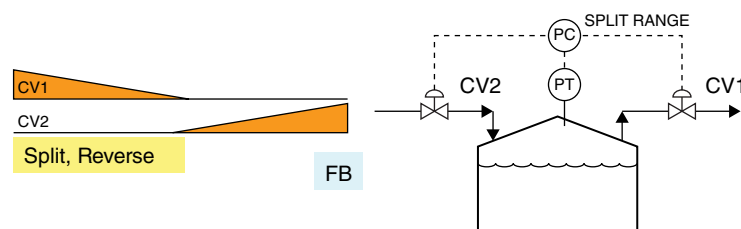
In Figure 14.29, we have blanket gas at the top of a tank and it is important to control the pressure through two control valves, CV1 on the inlet (blanket gas stream), and CV2 on the outlet (vapor stream). We know that the type of control used is split-range because it is written there. Some people just write 50% on the P&ID and then you know that it is split-range control.

Now, how do we work out if the two valves work in straight or reverse mode? We do this by analyzing the operation. If the pressure of the blanket gas in the vessel goes too high, we open CV1 to relieve the pressure and CV2 gradually closes until the pressure is stabilized at its set point. Then as the liquid level in the tank drops, the pressure starts dropping and CV2 will open gradually to increase the pressure again. So it is a reverse mode operation.

Table 14.10 Types of split range control.

	Straight	Reverse
Split		
Parallel		

Figure 14.29 Examples of split-range control – blanket gas.



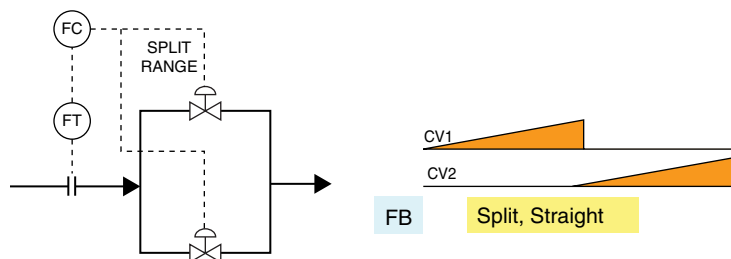


Figure 14.30 Examples of split-range control – wide range control valves.

This control system could work in parallel, but then with both valves open at the same time you would waste a lot of blanket gas.

If the mode of control cannot be specified easily, the process engineer needs to explain his intent.

The second example, shown in Figure 14.30, is of a wide-range control valve. Sometimes we have a lot of fluctuation in the flow in a pipe. The maximum flow rate can even be as high as 500 times the minimum flow rate. In a case like this, we install two control valves in parallel to regulate the flow with split-range control. When the flow rate is low, CV1 opens gradually to its maximum and then on a further increase in flow rate, CV2 will start to open. So this is the straight mode of split-range control, because the two valves are complementing each other. Sometimes this mode of operation is represented on the P&ID by writing 0–50 above one valve and 50–100 above the other. This indicates that the 0–50 valve is the first one in the sequence to open, and the other will follow.

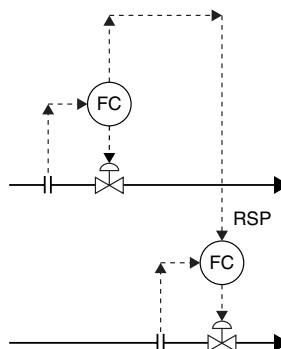
Table 14.11 shows the process cases that may call for the implementation of split- or parallel-range control.

Table 14.11 Process cases that call for the application of split- or parallel-range control.

Split-range control	Parallel control
Where two streams need to work together to set a parameter, and There is a preference in usage of one stream over the other stream	Where two streams should work together to set a parameter, and There is no preference in the usage of one stream over the other stream, and both streams need to work at the same time as complementary actions. One example is different units in parallel.

■ Cascade control:

A Controller signal from one loop goes as RSP to the other loop.



■ Ratio control:

A Sensor signal from one loop goes as RSP to the other control loop.

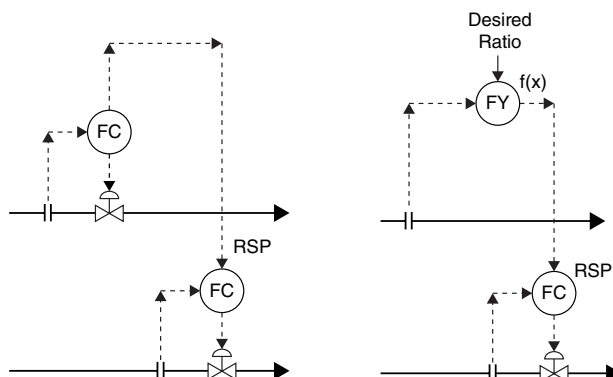


Figure 14.31 Cascade control versus ratio control.

The difference can be seen in Figure 14.32.

14.8.5 Clarification of Confusion

There are cases where two totally different loop structures “look” the same. Here I will try to show the difference to clarify the difference between the “similar-looking” cases.

14.8.5.1 Cascade Versus Ratio

Both of these functions are based on RSPs (remote set points) but in cascade control, a controller signal from one loop goes as an RSP to the other loop (and the two parameters have some relationship), while in ratio control, a sensor signal from one loop goes as an RSP to the other control loop. The other difference is that in cascade control two parameters are somehow interrelated.

The difference can be seen in Figure 14.31.

14.8.5.2 Single Loop Versus Ratio

In both of these, a *sensor signal* provides a set point (RSP) to a loop, while in ratio control, a *sensor signal* is “processed” and then sent as an RSP to the loop.

14.8.5.3 Selective Versus Override

In both of these, more than one signal goes to the selector function (high selector or low selector), but in selective control, the signals are *sensor signals*, while in override control, the signals are *controller signals*.

The difference can be seen Figure 14.33.

14.9 Monitoring Parameters

So far, our discussion has been about “automatic control.” However, not all parameters in plants are controlled automatically. Basically, the main types of control in each plant can be divided into two groups: “automatic control,” and “manual control.”

In automatic control, a system takes care of the control while in manual control, an operator collects the information from the plant and takes corrective actions based on that. The information that an operator needs for manual control can be obtained by reading the sensor and/or the data provided by the lab from the samples provided to them.

We like to control everything automatically, but there are some cases where we need manual control. There are cases where there is no “sensor” for a parameter, or the

available sensor is expensive. Another case is when a control task needs more judgment by a human. In such cases, we can also use manual control as long as the process parameter is not very “agile” and moves sluggishly (although these days, “expert systems” have been designed to be used in such cases). In such cases, we may decide to rely on manual control rather than automatic control.

For each piece of equipment, different process parameters (including pressure, temperature, flow rate, level, and composition) can be defined. However, not all of them are equally important.

All of the defined parameters for each piece of equipment need to be checked against the need for monitoring. The type of “monitoring” is determined based on the level of criticality of each parameter. If the parameter is critical, it should be automatically controlled. If it is not critical but very important, it can be visually checked in the control room through an indicator, and an operator can take action if the parameter is out of the normal band. If it is mildly important, it can be visually checked by the rounding (field) operator through a field indicator. And for a relatively unimportant parameter, there may be nothing, or only a “measuring point” somewhere in the plant for the rounding operator to use his portable indicator, to check the parameter occasionally.

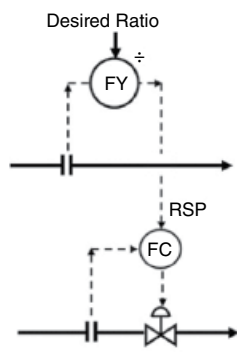
Therefore, for “very important” and “critical” parameters, automatic control is required, whereas other parameters are operator- or manually controlled.

This concept is shown in Table 14.12.

The decision on the level of criticality of each parameter depends on the type of equipment, the commodity type, the level of harshness of the environment, and the level of skillfulness of the operators. However, Table 14.13 can be used as a guideline.

■ Ratio control:

A Sensor signal from one loop goes as RSP to the other loop.



■ Single Loop control:

A Sensor signal goes to the controller.

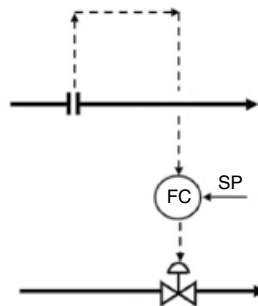
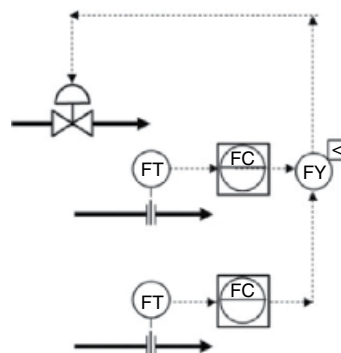


Figure 14.32 Single loop versus ratio control.

Figure 14.33 Selective versus override control.

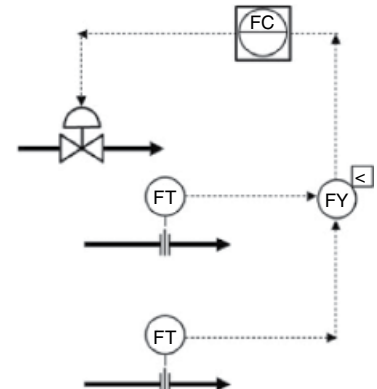
■ Override control:

Multiple controller signal from different location go to selector.



■ Selective control:

Multiple sensor signal from different locations go to selector.



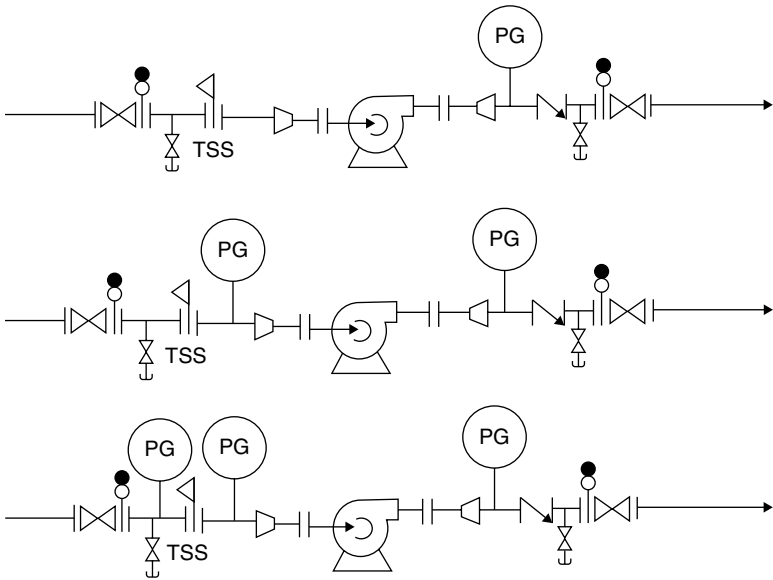


Figure 14.34 Pressure gauge location on a centrifugal pump.

Table 14.12 Deciding on the type of monitoring.

	Low priority	Medium priority	High priority	Critical
Required level of monitoring	Monitoring in the field using a portable gauge	Monitoring in the field	Monitoring from control room	Monitoring by control loop
Example for pressure as the parameter				

Table 14.13 Guidelines for deciding on monitoring type.

	Barely important	Mildly important	Very important	Critical
Arbitrary definition	Small or no effect over the long term	Long-term effect on product quality or smoothness of process operation	Quick effect on product quality or smoothness of process operation	The main purpose of the equipment, critical weakness of equipment

Here, we can see that not every instrument, gauge, sensor, or piece of equipment needs to be hooked into a control loop. I’m sure that you have seen instruments in a plant that are not tied in to any control system.

In Sections 14.9.1–14.9.2 monitoring of equipment is studied.

14.9.1 Container Sensors

Containers need a minimum level monitor, if they are non-flooded with liquids, or a pressure monitor if they contain gases or vapors.

Temperature could be another parameter for some containers. Temperature sensors are more common for

smaller containers and/or when the downstream equipment is very temperature sensitive.

14.9.2 Fluid Mover Sensors

The most important function of a centrifugal pump is increasing the pressure of a fluid stream. Generally the pressure of the fluid stream in the discharge of a centrifugal pump is controlled. Other than a control loop for the discharge pressure of centrifugal pump it is good practice to monitor the pressure of the discharge by the field operator. Therefore almost always there is a pressure gauge on the discharge side of a centrifugal pump.

In some cases another pressure gauge is installed on the suction side of the centrifugal pump. This could be the case when there is a high chance of cavitation in the pump. If the pumping liquid is very hot and/or very volatile there is a high potential for a cavitation phenomenon, which is detrimental for the pump. In such cases, it is a good idea to put another pressure gauge in the suction of the centrifugal pump to be monitored by a field operator and make sure that there is enough pressure in the suction of pump to prevent cavitation. Sometimes the temperature and pressure gauges are installed on the suction side of the centrifugal pump instead.

There are some cases that a third pressure gauge is placed around a centrifugal pump and that the pressure gauge is the one upstream of the strainer. If the pumping liquid is very dirty and there is a high chance of plugging the strainer it is a very good idea to see one pressure gauge upstream of the strainer and one other pressure gauge downstream of that. This provision helps the field operator to check these two pressures and make sure that the strainer is not plugged.

The case for PD pumps is not much different from centrifugal pumps. The only difference could be elimination of a pressure gauge on the suction side of the pumps as they are not sensitive toward cavitation.

14.9.3 Heat Exchanger Sensors

We know that almost always the outlet temperature of the target stream should be equipped with a temperature control loop.

In addition to this, all the streams around a heat exchanger are equipped with a temperature sensor and a pressure sensor. The reason for the temperature sensor is very obvious, as the main duty of a heat exchanger is changing the temperature. However, it could be questioned why there is a need for installing pressure gauges on a heat exchanger. The main reason for installing pressure gauges on the pipes around a heat exchanger is to identify if there is a leakage inside of the heat exchanger. As a heat exchanger consists of two enclosures, one for cold fluid, the other one for hot fluid, and are in contact with each other while they are not always visible, it is a good idea to monitor the pressure of the fluids to make sure no leakage or rupture happens. In the case of leakage or rupture the fluid migrates from the high pressure side to the low pressure side and the pressure raises the low pressure fluid.

Even though it was mentioned that pressure gauges and temperature gauges are installed on the streams around the heat exchanger, there are some exceptions and deviations.

For example, if the heat exchanger is a utility heat exchanger, it is not very common to see a temperature gauge on the utility stream. The reason is that the utility stream is intended for heat transfer. Its inlet temperature is fairly constant and the outlet temperature is not very important as it goes to a “temperature compensation system” such as a fired heater or an air cooler.

When two or more heat exchangers are stacked up on each other possibly no pressure gauge or temperature gauge are installed on the space between the two stacked heat exchangers. The reason is lack of room for these instruments. If it is really necessary to monitor the temperature and/or pressure of each single heat exchanger in a stacked arrangement a thermowell instead of temperature gauge and a pressure point instead of pressure gauge could be installed. A thermowell and a pressure point (PP) is installed and operators can use them to check the temperature and pressure of the stream using their portable temperature gauge and portable pressure gauge.

Sometimes, instead of installing a pressure gauge on the outlet side of a utility stream, only a PP is installed to save some money (Figure 14.35).

14.9.4 Fired Heater Sensors

Because of the criticality of fired heaters, they are equipped with a BPCS, SIS and monitoring systems.

Generally speaking all flows and all temperatures are monitored in a well-equipped fired heater.

On the process fluid side, the total flow and the flow of each pass are monitored.

On firing system flows and the flows of fuel and air are monitored.

On the flue gas side the pressure of the stack is monitored, which gives a sense of the flue gas flow rate too.

The temperature of the outlet process fluid, on each pass, needs to be monitored.

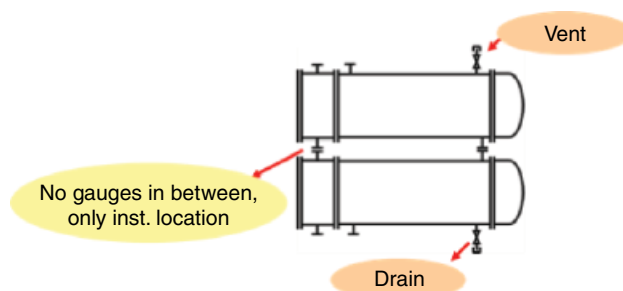


Figure 14.35 Monitoring of stacked heat exchangers.

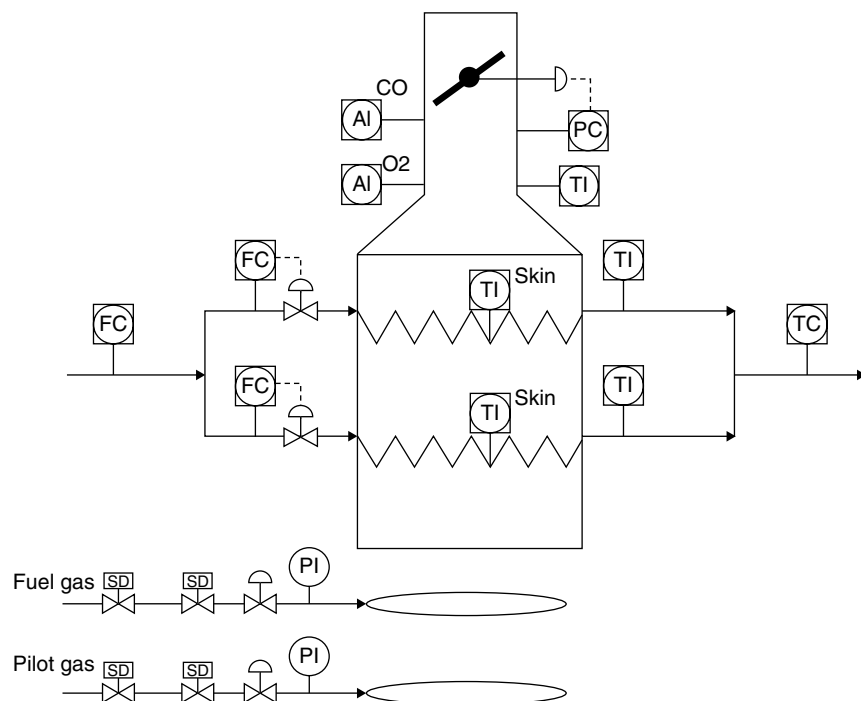


Figure 14.36 Monitoring of fired heaters.

The temperatures of the skin of tubes and also the fire box should be monitored.

If the fired heater is the area with stringent environmental regulations and/or the efficiency of firing is very important the flue gas components can be monitored by

process analyzers. The examples are carbon monoxide and the oxygen content of the flue gas.

A typical monitoring system around a fired heater is shown in Figure 14.36.

15

Plant Process Control

15.1 Introduction

What is the “plant control system,” and how we can implement it?

In more technical terms, how can we implement a BPCS (basic process control system) in a plant?

A BPCS, or regulatory control system, can be divided into two main levels in plants:

- 1) Plant-wide control. This is used to provide overall control for the entire plant. Some people refer to this as “heat and material balance control,” since its primary role is to ensure that the plant produces the product in the predicted quantity and with the predicted quality. This control basically creates a link between a plant and its H&MB (heat and mass balance) table. Not all processes and unit operations are individually and directly connected to the plant-wide control system.
- 2) Equipment control (unit operation control). Each “unit” within the process may need to be controlled via its own BPCS with corresponding control loop(s). Most pieces of equipment don’t have an operating “point”; rather, they have an operating “window”. This is not necessarily because of an inherent weakness of the equipment; this is something we like and gives the equipment the capability to “fluctuate” under different process conditions. The main duty of the unit control is to bring the unit to its optimum point within its operating window in each different set of process conditions.

It is important to mention that the classification of the BPCS control into two levels is only based on their concepts; generally there is no difference in control hardware in a plant, and these two groups cannot be recognized or differentiated in P&IDs.

Each of the above concepts carries one aspect of plant control. Plant-wide control assures the “attachment” of the plant to its capacity and the quality of product(s), whereas equipment-level control tries to bring a piece

of equipment to its best operating point within its operating window and also protect the equipment at its weak points.

While in a P&ID all control systems can be traced, a PFD generally shows only plant-wide control. However, some P&IDs show some major elements of equipment-level control too.

Plant-wide control is discussed in Sections 15.2–15.4 of this chapter while equipment-wise control will be discussed in Section 15.5 and after.

15.2 Plant-Wide Control

There are two purposes to installing plant-wide control:

- 1) The main purpose is to link the plant to its heat and material balance table.
- 2) The second purpose is surge or disturbance management.

15.3 Heat and Mass Balance Control

The first purpose of plant-wide control is heat and material control.

Some people may prefer using the phrase of “mass balance” or “material balance” control (rather than “heat and material balance control”). Their logic is: “in plants, we care about flow rate and quality of the product(s); no one is looking for a specific product with a specific temperature.” While their logic is true in the majority of cases, it should be noted that to have a product with a specific quality, it may need to go through different steps of operation, which need specific operating temperatures. Therefore, “heat control” is still needed.

In order to ensure this, theoretically we must have at least one flow control loop and one composition control loop with a manual set point that comes from the H&MB table.

However, we don't always see composition control as a part of plant-wide control. The reason for this is that we generally don't like to use "composition" control loops, and instead we use a control loop with a parameter that "infers" the composition, with a manual set point. We don't really like composition control loops because their "sensors," or more technically speaking, their process analyzers, are not very reliable instruments, and they are also expensive.

As stated previously, in a plant, control loops have different types of set points: manual set points and remote set points (RSP). We like RSPs since they help to operate a plant smoothly. However, it is essential to have at least one manual flow set point to act as a "policeman" and ensure that the process conforms to its material balance.

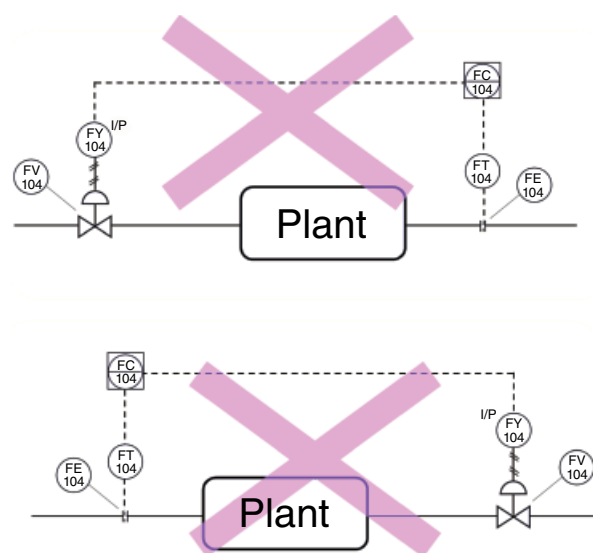


Figure 15.1 Plant-wide control – first attempt.

However, because flow control loops with manual set points complicate the smooth operation of a plant, we try to limit them to the "border locations" of the plant, rather than middle points.

"Border locations" means at the beginning, near raw material pipes, or at the end, on end product pipes.

Now the question is: how to do mass control? The simplest answer is by placing a control loop (as below) to control the flow rate in the whole plant. There are two available options and both of them are shown on Figure 15.1.

Neither of these works properly for plant-wide mass control, for different reasons. One reason is the slowness of the loop. Such a big loop is very sluggish. The other problem is that a plant is not a closed loop, from the viewpoint of the individual streams. There are multiple streams that tie into or branch off from the main streams; therefore, a "global single control loop" doesn't work properly.

Such a "global single control loop" may work in closed-loop plants (similar to some systems in HVAC industries), or when the plant is small.

So, how can we solve the problem? The solution is using multiple small control loops instead (Figure 15.2).

However, we don't like to see flow loops with manual set points, except at the "edge" of the plant; also, incorporating RSPs for flow loops forces us to use complicated multi-loop control architectures, like cascade control, which we don't like.

Here we have to use more diverse types of control loops.

Mass balance control is nothing more than the interaction between inventory and flow. Basically, the proper interaction between the inventory of fluid inside a plant, and the flows in the pipes in a predetermined way, guarantees the mass balance and the required plant-wide hydraulics.

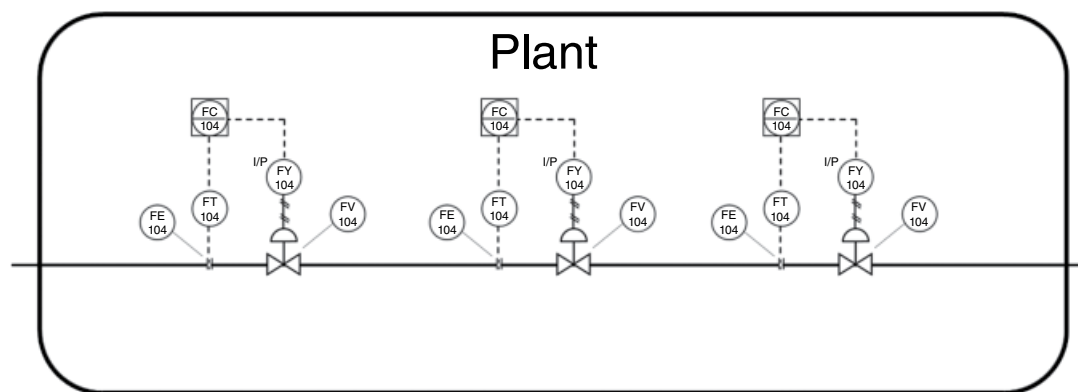


Figure 15.2 Plant-wide control – second attempt.

Table 15.1 Plant hydraulic control.

	Purpose	Gas/vapor	Liquid
Container	Inventory control	P-loop	Non-flooded: L-loop Flooded: F-loop
Pipe	Material transportation control	P-loop (or F-loop if it is around a gas mover)	F-loop

Here, plant-wide hydraulics means the adjustment of parameters in such a way as to guarantee the movement of fluid from point “A” at the beginning of a plant to point “Z” at the end of the plant.

Flow, pressure, and level control loops are what primarily dictate the hydraulics of a process plant.

Table 15.1 shows their functions in different process items.

This table basically says:

- A pressure loop in a gas container is similar to a level loop in a non-flooded liquid container.
- A pressure loop on a gas pipe is similar to a flow loop on a liquid pipe

However, it should be mentioned that it is not very common to use pressure loops on liquid-filled enclosures (including pipes and containers).

Possibly the best approach is to use flow/pressure loops paired with level/pressure loops on an adjacent container (Figure 15.3).

However, the question still remains: what would be the “arrangement” of flow/pressure loops paired with level/pressure loops on an adjacent container? This question will be answered when we learn about surge control.

15.4 Surge Control

The second purpose of plant-wide control is surge management and control.

Change, surge, disturbance, fluctuation, or whatever you want to call it, is part of our life. Everything around us is changing. However, a plant is supposed to generate a product with a specific flow rate and a specific composition (quality). Therefore, somehow surge needs to be managed in a plant. This is one of the duties of plant-wide control.

Be careful: when we install a control valve on a pipe, we don’t “eliminate” surge; we only try to prevent the surge from spreading to downstream equipment. This “blocked” surge needs to be managed somehow.

The following section explains different types of disturbances in a process plant.

15.4.1 Disturbances in Process Parameters

Let’s have a look at some examples of disturbances occurring with each process parameter:

- Flow. Process plants always experience disturbances in flow. This is the most important and most frequent parameter to consider when considering disturbance management.
- Temperature. You can have a disturbance when there is a change in ambient temperature that will affect the process. Temperature disturbance can be compensated for (managed) in heat transfer equipment. There are two main types of heat transfer equipment: fired heaters and heat exchangers. Temperature disturbance is managed in a better way if there is a fired heater. When there is a heat exchanger, the temperature disturbance can be managed in a better way when it is a utility heat exchanger, rather than a process heat exchanger.

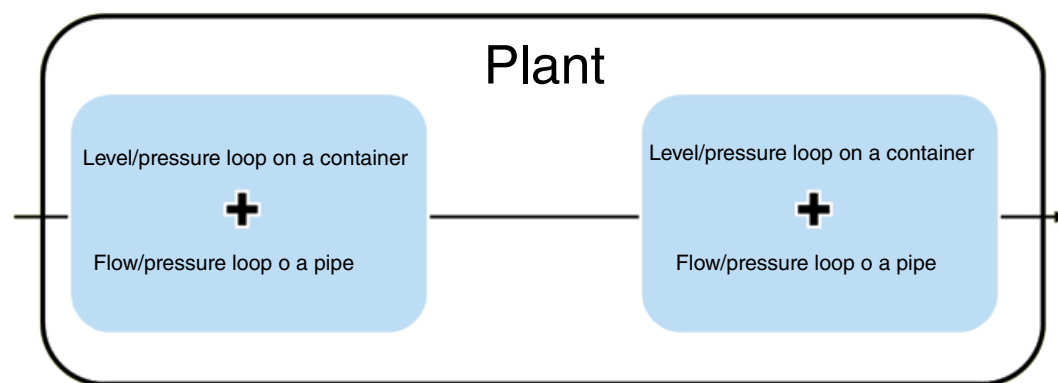


Figure 15.3 Plant-wide control – third attempt.

- Pressure. A change in flow can cause a change in pressure. In a compressor, a change in density of a gas will cause a corresponding change in pressure.
- Level. This can change when there is a change in flow rate.
- Composition. It's not very likely that a change in composition will cause a disturbance of significant magnitude in a process. If it does, it is very slow acting.

It can be seen that surge management can be summarized into two types of fluctuations: flow fluctuation and temperature fluctuations. The reason for this is that by resolving a flow surge, pressure surges and level surges are also resolved in the majority of cases.

Normally we use level loops to control surges when dealing with liquids, and pressure loops for gas surges. Small composition fluctuations can also be resolved by placing containers along the route, which is one form of flow surge mitigation.

15.4.2 Disturbance Management

There are two main ways to manage a disturbance in a process:

- 1) Absorption of surge
- 2) Rejection of surge.

15.4.2.1 Absorption

Absorption means absorbing a surge within the “process” and in the equipment in the plant. This disturbance-handling method is specific for flow surges. There is practically no such solution for temperature surges; however, by implementing an absorption method to dampen a flow surge, temperature surge is partially dampened. There are three ways we can achieve this:

- 1) Spreading it out over multiple pieces of equipment. In this method, we pick a few pieces of equipment over which to spread and dampen the surge. First, we have to ensure that each piece of equipment is able to tolerate its share of the disturbance within its design parameters. Then, by placing control loops, we “direct” the surge into the selected piece of equipment. In this method, we generally pick multiple vessels over which to spread the surge.
- 2) Dampen it in a “buffer” (a robust system). In this method, we pick only one piece of equipment to box in the flow surge, and dampen the surge in it. The classic example of a robust or buffer system is a surge container or tank.
- 3) Condensation. If/when the surging flow is vapor, condensation could be a good solution to dampen the surge. In this method, we remove the “picks” of the

surge diagram by condensing a portion of vapor to a liquid and decreasing the volume and then pressure.

Management of surge through absorption will be discussed in more detail in Sections 15.4.4 and 15.4.5.

15.4.2.2 Rejection

This method of dealing with a disturbance involves rejecting it from the process to outside of the “process”. This method can be used for flow surges. It is a procedure that means losing either fluid or gas to waste. We don't like to do this, because it can mean the loss of valuable process fluid that cannot be regained.

Some examples of rejection include venting a gas or vapor to atmosphere, sending fluid to a waste line, and sending gas to burn in a flare system. Rejection of a disturbance is the last line of defense in a process control system (or BPCS); we generally prefer to use rejection in SIS functions rather than in BPCS loops.

Another example is rejection of temperature surge to a heat transfer utility network. In a utility heat exchanger, where one stream is process and the other is a utility stream, you can manage any surges in temperature in the process stream by transferring those fluctuations to the utility stream. So here, the temperature disturbances in the process stream are rejected to the utility stream.

While we may not be concerned about transferring temperature disturbances to a utility stream, the same cannot be said for a process stream. So we would have a problem managing disturbances in a process heat exchanger where we have two process streams exchanging heat. Obviously one of the process streams must absorb the temperature surge, which is not what we want. How do we solve this?

This is solved by having two heat exchangers in series; first a process heat exchanger, followed by a utility heat exchanger. In this case, the first heat exchanger is called “rough,” e.g. a rough heater or rough cooler. The second one is called a “trim” heat exchanger, e.g. a trim cooler or trim heater. It is important to remember that the last heat exchanger in the series is always a utility heat exchanger. Therefore, if good control/disturbance management is needed for temperature, two heat exchangers in series are used. Some people say that trim heat exchangers always have a duty of one-third of its corresponding rough heat exchanger, but this is not always the case.

15.4.3 Disturbance Versus Fluid Phase

The way we manage a disturbance could be different depending on the type of fluid phase (liquid or gas/vapor).

We learned that disturbance can be managed by two main methods: absorption or rejection. However, for

Table 15.2 Disturbance control of different fluids.

	Liquid	Gas/vapor
Absorption	Yes Requires spreading the surge over different units. Requires a surge container.	Yes Requires spreading the surge over different units. Generally no surge container is used.
Rejection Generally the last action to deal with surge	Not very common Unless a plant is handling unpredictable liquid surges, a liquid pond can be provided as the rejection destination.	Very common The rejection destination could be atmosphere or flare.
Cooling/condensation	Not effective for liquids	Generally the first action to deal with surge for gases/vapors. Cooling: works for gases and vapors Condensation: works for vapors.

gases there is one other option available: this is decreasing the volume of gas by cooling it or condensing it (if it is a vapor).

A disturbance can basically be managed by placing a control valve on an appropriate piece of equipment. However, by doing this, the surge is – partially – stopped, and will report somewhere upstream of the control valve. Depending on whether the fluid is a gas or vapor, this mitigated disturbance can be managed in different ways.

These methods are summarized in Table 15.2.

Generally speaking, when the size of the surge is large, it should be handled by rejection; absorption doesn't dampen the surge in an effective way. However, the word “large” here has a different meaning if the surge is liquid or gas. The threshold for “large” for gas surges is much lower than the threshold for “large” in liquid surges. That is why we see more pipes connected to flare in gas processing plants than pipes connected to a liquid pond (if any) in liquid processing plants.

When it comes to managing a surge by absorption, again there is a difference between gas surges and liquid surges. Liquid surges need a “surge container” in the majority of cases, while gas surges can be handled in the existing string of units, possibly without the help of a surge container.

The reason for this is that although containers or voluminous equipment in liquid services cannot handle liquid surges very well, dedicated surge containers in liquid services can do so.

This concept is the reverse for gas surges.

In gas services, containers or voluminous equipment can handle fairly large gas surges very well (without ignoring their specific duties), but if a dedicated surge container is placed in gas service, it cannot handle large gas surges very well.

That is why surge vessels in gas processing plants are not as common as surge tanks in liquid processing plants.

15.4.4 Dampening Gas/Vapor Flow Surge

The image below shows a few examples of handling gas/vapor surges (or pressure control) through different methods.

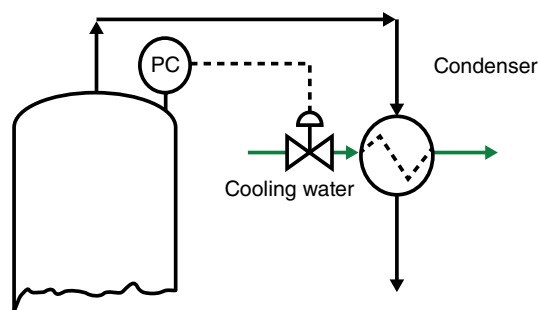
In some documents, instead of using the word “gas,” the phrase “non-condensable gas” is used instead.

Figure 15.4 shows the first example where the pressure of a vessel (here, a distillation tower) is adjusted by condensing vapor. When the pressure goes high, the control valve will open to send more cooling water to condense more vapor and consequently the pressure of vessel will decrease. However, the method doesn't work properly if there is a significant amount of non-condensables in the vapor stream.

Here you want to dampen the pressure surge in the vessel. We use a condenser to transfer any pressure surge in the tank.

However, if surges are large, then the rejection solution should be used.

Figure 15.5 shows a pressure sensor mounted on the top of a distillation tower, but it could be installed on any part of the vessel whose pressure you want to control. Here we have a pressure sensor going to a pressure controller, which activates a control valve on an outlet line.

**Figure 15.4** Surge dampening through condensation.

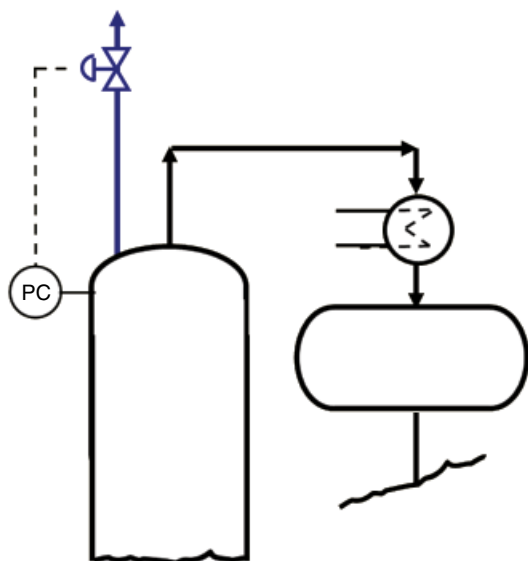


Figure 15.5 Pressure loop for large surges.

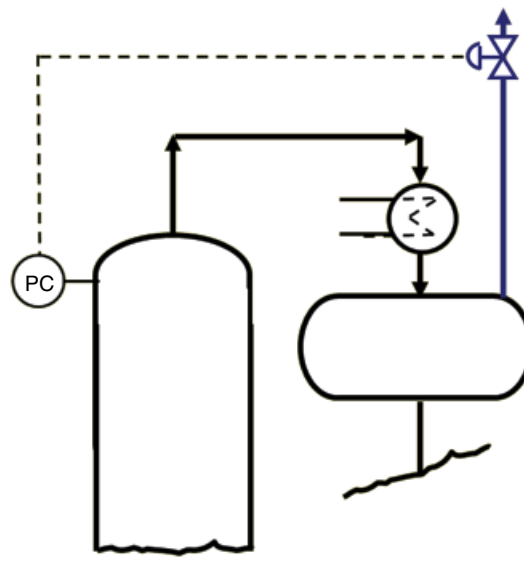


Figure 15.6 Pressure loop for a large amount of vapor.

In the case of flammable vapors, this outlet line could go to a flare, or it could vent to the atmosphere in the case of a harmless vapor like steam.

However, this solution wastes a large amount of valuable fluids. If there is a significant amount of vapor – rather than gas – the same methodology can be used, but gas release could be from the drum after the condenser (Figure 15.6).

15.4.5 Dampening Liquid Flow Surge

As was discussed, managing surge through absorption is the preferred way of handling surges in a plant. The absorption of a surge is basically done by spreading the surge around, and then dampening the residual distributed surge.

Doesn't this ring a bell? Yes, we already said “flow + inventory” control can be done by a pair of flow/pressure loops on a pipe and level/pressure loops on an adjacent container. Therefore, spreading + dampening can be

done by pairing flow/pressure loops on a pipe and level/pressure loops on an adjacent container.

But the question that remained unanswered was: “what is the best pair?” Option (a) in Figure 15.7 or option (b)?

As can be seen, there two available options:

- The upstream option, where every level loop takes its level signal from an “upstream” container.
- The downstream option, where every level loop takes its level signal from a “downstream” container.

But which one is used in which case? To be able to answer this question, we need to know the capability of each of the options or strategies. Both of them are capable of spreading out the surge, but in different ways:

- The upstream strategy works best for spreading out surges coming from upstream.
- The downstream strategy works best for spreading out surges coming from downstream.

These are shown in Figures 15.8 and 15.9.

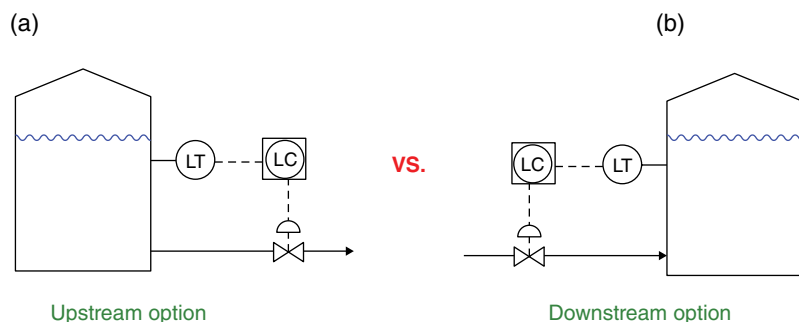


Figure 15.7 Two options of level + control valve pairs.

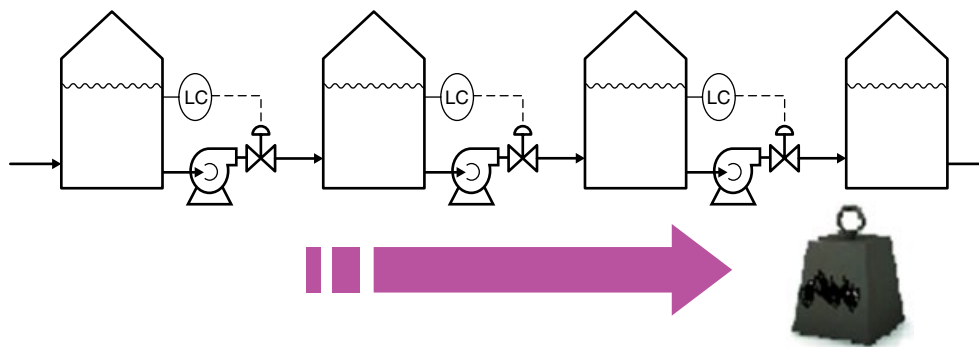


Figure 15.8 Scheme one – spreading a surge: upstream strategy.

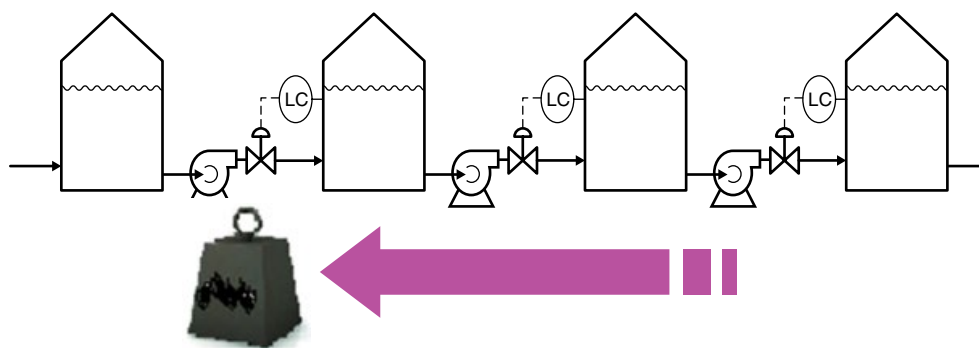


Figure 15.9 Spreading a surge: downstream strategy.

This scheme can be named the “upstream strategy” because each control valve receives an order from the “upstream” container. This strategy is good and blocks the fluctuation when it occurs upstream and it propagates toward downstream.

Scheme 1 is more popular for process streams in industry because that is where most disturbances originate.

Scheme 2 shows the downstream strategy of control, where each control valve takes action on a signal from the downstream tank. This arrangement is good when a disturbance spreads from downstream toward upstream. In such cases the downstream strategy blocks the surge and prevents it from going upstream.

This downstream strategy is the one most used for disturbance management in utility streams.

By now, you can start to recognize that control systems in liquid processing plants rely heavily on containers. This means that containers may have control duties on top of process duties. There are even a few cases where we add a container purely for the purpose of control.

One example is a “condensate pot,” which we will talk about in Section 15.8.1.3.

If we design a process with a minimum number of tanks and just move the product stream from one unit operation to the next, it makes it very difficult to control. A control engineer would like to see tanks in the process to manage surges and make his job far easier.

Let’s have a look at some fundamental examples of containers in plant-wide control.

The first example is a pure surge tank. Figure 15.10 shows that as there is a flow loop with a manual set point, the surge is almost completely damped in the tank. This tank blocks the surge fairly well, but small surges may escape from the tank and ripple downstream.

The other downside of this arrangement is that it limits the flexibility of operation because of the manual set point on the flow loop.

This tank cannot be used as the feed tank for a critical item. The main duty of a feed tank is to make sure there is enough feed for the downstream item in the near term. Here, because the arrangement is tailored to make sure

there is a specific flow rate downstream, no one can guarantee that there is always enough liquid in the tank, so it cannot be used as a feed tank.

In Figure 15.10, a pure surge tank is shown; this should generally be a large tank (probably not a vessel), because you have to rely on the tank volume to be able to block surges.

In the next example, we have tried to design an arrangement with a higher level of surge dampening capability. In this arrangement, we use a “downstream strategy” that effectively dampens surges from upstream. And then to make it a better design, we have added a flow loop control downstream of the tank to make sure no surge can escape from the system (Figure 15.11). Here we still have the downside of the lack of flexibility in operation because of the manual set point on the flow loop.

The container size in this example could be smaller than the example one. This is because we are controlling the flow to the tank on the feed line in this example.

The next example is a pure feed tank. In this example, we try to keep a specific level in the container. In this arrangement, we use an “upstream control strategy,” so surges coming from upstream are not blocked. Therefore, no surge dampening capability exists here. This arrangement is used when there are no large surges predicted, or the surges are already dampened upstream (Figure 15.12).

As stated above, the arrangement has to have the capability of surge dampening, so there is no guarantee of a fixed flow downstream of the vessel. However, using a trick we can add a mild capability of surge dampening to the

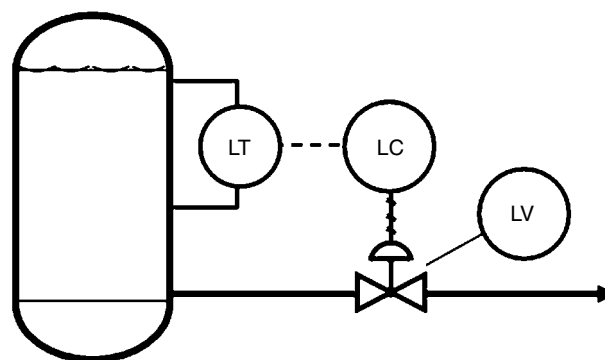


Figure 15.12 Pure feed container.

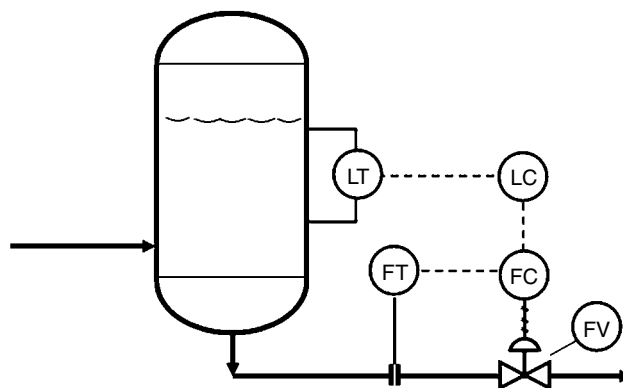


Figure 15.13 Feed container plus mild surge dampening capability.

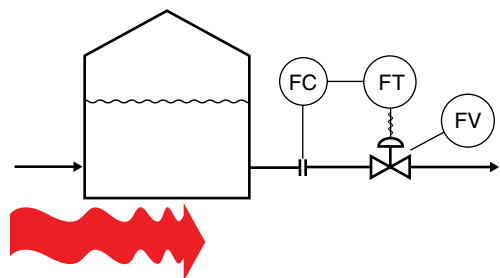


Figure 15.10 Pure surge dampener.

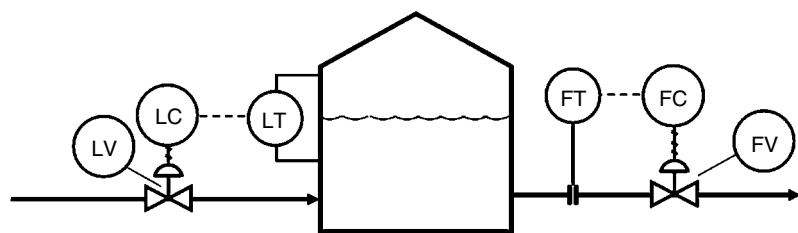


Figure 15.11 Surge dampening using a tank to block + transfer surge.

above arrangement. This is the trick we learned in Chapter 14. If we keep the vessel size small enough, by applying cascade control on an already tight parameter, we may control two parameters in a cascade control arrangement fairly well. Figure 15.13 shows this arrangement.

The vessel in Figure 15.13 could be a feed tank, which could be upstream of a distillation tower.

The next example is the most expensive arrangement; it works to dampen the surge as well as working as a feed tank.

The arrangement shown in Figure 15.14 exhibits perfect disturbance management but it is very expensive.

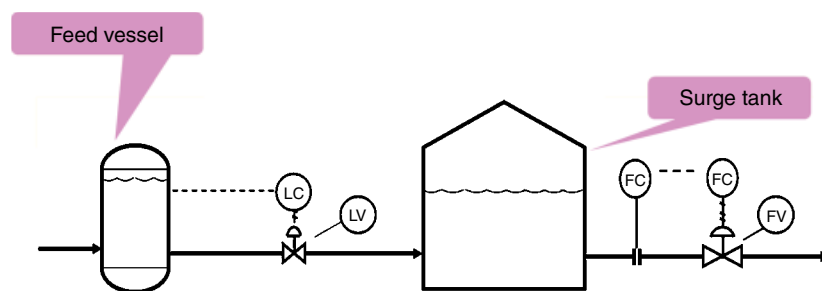


Figure 15.14 Surge dampening using feed + surge tanks.

It is ideal for critical items of equipment downstream, like a critical fired heater (for example a reactor fired heater), where you need a perfectly controlled flow rate on the feed line. In order to achieve this, we have two tanks in series: one operating as a feed tank and the next as a surge tank. In previous examples, we had combined the feed control and surge management duties into one tank to save money.

15.4.6 The Purpose of Containers in Process Plants

From the above discussion, you may recognize that there are at least five different roles for containers in process industries. They are:

- 1) Containers for physical or chemical conversion. These vessels are designed to perform a specific process goal and are often referred to as unit operations, or process units. This type of vessel can be a reactor or a separator. The separator can be a two-, three-, or even multi-stage operation such as a distillation column.

- 2) Containers for long-term storage of material. Feed storage tanks or product storage tanks are examples of these. They are generally huge tanks.
- 3) Containers for medium-term storage of material. This could be a day tank, designed with a holding capacity in order to buy time for maintenance of upstream units with the minimum interference on the downstream units.
- 4) Containers for dampening surges. These vessels are used to dampen a surge by blocking the surge. We call these surge tanks or equalization tanks.
- 5) Containers for assuring continuity of flow. These vessels are named feed tanks, and their function is to wipe out short-term discontinuities in flow for the benefit of the downstream unit. Not all units need such feed tanks. Feed tanks are used when the consequence of a sudden drop in flow is not tolerable. For example, if a distillation tower sees a quick change in its feed flow, it may take up to an hour to stabilize.

The different types of containers are summarized in Table 15.3.

Table 15.3 Container roles in process plants.

Name	Different names	Storage tank	Day tank	Surge tank (vessel)	Feed tank
Examples	Reactor, separator, distillation tower,	Crude oil storage tank, milk balance tank, etc.	Tanks at the end of each major unit	Wastewater equalization tank	Fired heater feed tank
Goal	Physical or chemical conversion	Storing material	Minimizing the impact of maintenance on a set of downstream units	Smoothness of operation and control	Assuring the continuity of flow
Duty	Converting the material physically or chemically	Storing the material for long term	Storing the material for medium term	Providing fairly fixed flow rate to downstream units, dampen surges	Storing the material for a short term, provide flow continuity in the near term to downstream units
Where	For process purposes	At the beginning and/or end of a plant	Between each set of units	In the middle of a process string	Upstream of some items like furnace, distillation tower
Typical residence time	Few minutes or less	Few days	8–24 hours	Less than half an hour	Fraction of an hour

It is important to note that some tanks have a dual (or even triple) purpose, like a day tank that also functions as a surge tank. So, if at some point it is no longer necessary for one purpose, it may still be vital for the other. Process and control engineers have to take this into account when optimizing equipment in a process plant.

15.5 Equipment Control

This topic has three main sections. In the first section, we will learn whether we need to control a specific piece of equipment at all. In the second section, we will learn the main ideas for creating a control system around a piece of equipment (if the answer to the first question is yes). In the third and last section, we will learn about a general control system for common process elements.

15.5.1 Do We Need to Control at All?

At the beginning of this section, we should ask ourselves the following question: “do we need to control a system at all?”

We need to control a piece of equipment when:

- 1) A disturbance exists
- 2) The disturbance is not tolerable
- 3) The system is not self-controlled.

These requirements actually force us to put control on most equipment. The first feature is almost always present, since we live in a world full of changes. A disturbance is almost always not tolerable, except for in the case of a few process elements. For example, level disturbance could be tolerable for big tank, but possibly not for small vessels. Process items are almost always non-self-regulated.

15.5.2 Principles of Equipment-wise Control

To control each piece of equipment, a control loop, or control loops, need to be installed to satisfy the two requirements below. The result of addressing the requirements below could be the requirement to install one, two, three, or more control loops around the equipment.

- 1) Placing the control loop on the parameter which is the main purpose of the equipment. This is self-explanatory. If you have a heat exchanger, you place a temperature control loop on the pipe of the target stream. If you're dealing with a pump, you install a flow control loop. If you have a compressor, you need to install a pressure control loop, although sometimes we prefer to install flow control; these two are inter-related. With a distillation column or a reactor, the whole purpose of

Table 15.4 Parameter of the main purpose for different equipment.

Process element	Parameter to be controlled
<ul style="list-style-type: none"> • Tanks • (Liquid) non-flooded vessels 	Level
<ul style="list-style-type: none"> • Gas vessels • (Liquid) flooded vessels 	Pressure
<ul style="list-style-type: none"> • Pumps • Compressors • Fans • Blowers 	Flow rate or discharge pressure
<ul style="list-style-type: none"> • Heat exchanger • Furnaces or boilers 	Temperature of target stream
<ul style="list-style-type: none"> • Unit operations • Unit processes 	Composition of discharge stream, or preferably “inferred” parameters plus Container control system (refer to tank or vessel control in this table)

the equipment is to change composition, so you could install composition control loops, although we prefer to avoid composition control loops.

A control loop based on this requirement brings a piece of equipment onto its intended operating point within its “operating window.”

Recognizing the “main process purpose parameter” is not difficult; Table 15.4 lists the most common pieces of equipment and their “main process parameter” that needs to be controlled.

- 2) Placing the control loop on other parameters and/or other streams' parameters to bring the equipment into its “operating window.” We generally don't install a control loop for every single parameter (temperature, pressure, level, flow, and composition), nor on both ends (high and low) to make sure that the unit operates within its operating window. We only install control on the “weak point” of the equipment if the risk of reaching that weak point is high. “Weak point” here means a process parameter related to the equipment, whereby if a process stream reaches that point there is a chance of a process upset or of losing the equipment due to equipment failure. If the risk of such a condition is low, you may decide not to install any control loop for this purpose. If the risk of such a condition is very high, you may decide to leave this duty to the SIS (safety instrumented system), rather than the BPCS, and not install any control loop for this purpose.

When talking about risk, we know that risk is a multiplication of “probability” and “severity of consequence”. For example, we need to decide if a pump needs a control

Table 15.5 Example of the minimum flow requirement for pumps.

	Explanation of case	Probability of flow change	Severity of consequence	Risk	Action
Cases 1	A pump in a closed loop	Nil, because it is a closed loop with constant flow	Loss of pump	Zero	No control loop to protect the pump against low flow
Case 2	A small pump on an (open loop) pipe	High chance	Loss of pump, but pump is small and inexpensive so the severity is almost nil	Almost zero	No control loop to protect the pump against low flow
Case 3	A large pump on an (open loop) pipe	High chance	Loss of large, expensive pump	Not negligible	Needs a control loop to protect the pump against low flow

Table 15.6 Parameters for “operating window” control.

	High	Low
Flow	Generally no need for control except for very sensitive units	Needs control when the unit is sensitive to low flow
Level	Generally addressed in SIS actions or override control (or BPCS and override control)	Generally no need for control unless specific submergence is needed for an element in the container
Temperature	Generally addressed in SIS actions	Generally addressed in SIS actions
Pressure	Sometimes controlled and sometimes handled by SIS actions	Rare to install control or SIS
Composition	Generally addressed in SIS actions or override control	Generally addressed in SIS actions or override control

loop or not, based on the “weak point” concept. Generally, pump manufacturers give a warning of damage to the pump if the flow rate drops below a certain level. Now let’s check the risk of this event for the three different cases in Table 15.5.

The three examples above show that the statement that “all pumps need to be protected from minimum flow by a control loop” is not set in stone.

Table 15.6 shows the parameters that may need to be controlled as part of operating window. However, it should be stressed that here there is a large amount of room for judgment by the designer.

Another example would be to control both the inlet and outlet pressure to make sure pressures stay within an acceptable range. Usually this is not necessary for every piece of equipment in the plant, but there are instances where equipment is sensitive to pressure, e.g. a process analyzer. Sometimes we replace a pressure control loop with a pressure regulator (Chapter 7).

Capacity Control: Yes or No?

Do we need to put a flow control loop on the inlet to every single piece of equipment? Isn’t it true that every single piece of equipment has a specific capacity and its flow rate needs to be controlled?

Actually the answer is no! We don’t install a flow control loop for every single piece of equipment even though we know each piece of equipment has a specific capacity. There may be different reasons for this, but one reason is that units are not very sensitive to the flow rate, and they can handle a higher or lower flow rate in cases of mild process upsets and short-term consequences.

The other reason is that we place a “feed container” upstream of flow-sensitive units instead of trying to deal with flow surge in the unit.

The rest of this chapter provides an overview of the general control loops around the most common process elements. We will cover control loops for pipes, fluid movers (pumps and compressors), heat transfer equipment (heat exchangers and furnaces), and containers.

You may be surprised to see “pipe control” in this section, since a pipe is not a piece of equipment. Even though this is the case, the different control procedures for pipes are very popular in process industries and they should be learned. That is why I used the phrase “process elements,” rather than “process equipment,” at the beginning of this section.

15.6 Pipe Control System

There are three areas of control to consider for pipes:

- 1) Control on a single pipe. All you need to do is decide if you're going to control the flow rate or pressure in the pipe.
- 2) Dividing the flow between multiple branches. We can also call this "branching off". We have the option of sending part of the flow to two or more users simultaneously, or non-simultaneously. The first case needs a control system but the second, which can be referred to as "diverting" a flow, can be done more easily. For flow diversion, several manual or automatic isolation valves can work together to divert a flow from user A to user B. (The other option is using one or a few multiport valves to divert the flow to a preferred destination. However, multiport valves are not always available in bigger sizes, i.e. larger than 8 to 10 in. NPS).
- 3) Merging flow from multiple branches. We call this "tying in". This mechanism is similar to dividing the flow. We can have two or more branches merging simultaneously, or we can once again use an arrangement of multiple isolation valves, or a multiport valve, to merge or introduce another stream into the main process stream from a preferred source (preferred source will be discussed in Section 15.6.2.1).

Below, we will discuss each of the above scenarios one by one. For dividing and merging flows, we will only focus on the simultaneous function, which is a more complicated case than the non-simultaneous function.

15.6.1 Control of a Single Pipe

It's important to understand what happens when you activate a control valve. When the valve starts to close, it generates more pressure drop. This causes extra pressure drop along the route, making the route less attractive for flow, so the flow rate decreases too. This means that by partially closing a control valve, the pressure on the downstream side of the control valve is decreased and the flow of the pipe will be lower as well.

It is important to know that, even though we said that a control valve on a pipe will decrease pressure and flow, no control valve can drop the pressure to a desired level and, at the same time, drop the flow rate to a desired level. We can adjust a control valve based on only one of these parameters. We can drop the pressure to a pre-determined level, but the amount that the flow rate drops is beyond our control. Likewise, we can drop the flow rate, but the amount by which the pressure decreases is beyond our control.

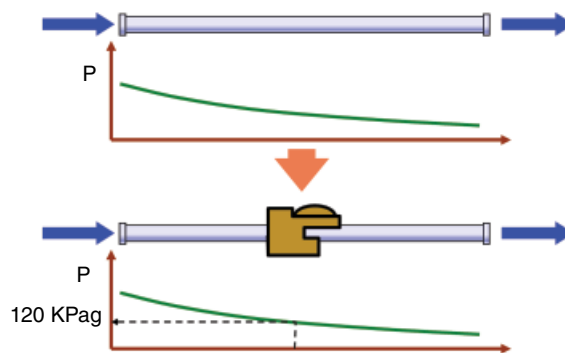


Figure 15.15 Pressure change in a pipe.

Where do we place a control valve on a pipe? There is only one rule to remember: if pressure is the parameter that you need to control, you place the control valve wherever the pressure is changing. The same applies to flow; if this is the important parameter, then you place the valve wherever the flow is changing.

15.6.1.1 Control of Pressure in a Pipe

The top schematic of Figure 15.15 shows a trend of pressure change within a pipe, where the stream is flowing from left to right. It is clear that pressure is decreasing at a rate with a specific slope. This slope is dictated by the pipe's friction and the types and number of fittings on it.

Now assume that we place a particular piece of equipment somewhere around the middle of this pipe (I intentionally used an irregular symbol in order not to refer to any specific equipment). The inlet pressure of the equipment is dictated by the pressure drop along the pipe, including the newly added equipment. Therefore, this pressure (which is 120 kPag here) is out of our control.

- What if this equipment needs 150 kPag as its inlet pressure? In this case, we need to put something at the beginning of the pipe, like a pump, to increase the pressure.
- What if this equipment needs 100 kPag as its inlet pressure? In this case, we need to put a control valve ahead (upstream) of the equipment in order to drop the pressure to 100 kPag (from the current pressure of 120 kPag).
- Can't we place a control valve downstream of the equipment to adjust the pressure? Yes, we can! As long as the pressure sensor of the loop is on the equipment side (which means upstream of control valve), the system will work fine. Figure 15.16 shows two available options for controlling pressure at one point along a pipe.

15.6.1.2 Control of Flow in a Pipe

The first question that should be answered is: do we need to control flow or not? There are actually several cases

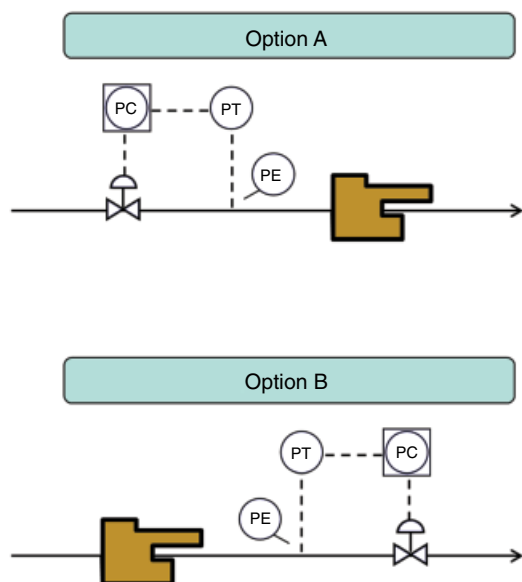


Figure 15.16 Control of pressure in a pipe.

where we don't really need to control a flow. For example, a stream that goes to a large storage tank possibly doesn't need to be flow controlled; an intermittent flow may not need to be flow controlled either.

If the flow of a stream needs to be controlled then the second question is how?

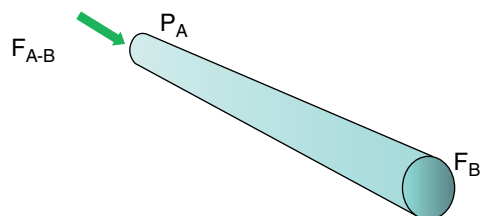


Figure 15.17 Control of flow.

Flow in a pipe is a function of the pressure at both ends, P_A and P_B .

In the schematic in Figure 15.17, if we increase the pressure at point A the flow (F_{A-B}) will increase; if we increase the pressure at point B the flow will decrease. This means the flow rate, F_{A-B} , doesn't stay at the value specified in the mass balance table.

The flow rate is actually a function of pressure at points A and B.

Therefore, if you try to fix the flow rate in a pipe, you need two pressure loops at either end (Figure 15.18).

Here what we are trying to say is if the intention is to control flow in a piece of pipe, two control valves are needed; one at each end of the pipe!

Even though this is theoretically a requirement, in reality there are not many cases where we need to install two control valves to be able to control the flow rate. We only need to install two control valves if the pressures at both ends of the pipe are fluctuating. If somehow we can prove that in one specific case the pressure on one side of a pipe is fairly constant, we may be able to get rid of the control valve at that "fairly fixed pressure" point and use a single control valve arrangement.

Table 15.7 shows the points where we can assume a fixed pressure, and points where the pressure is most likely fluctuating.

Table 15.7 Fixed pressure points and fluctuating pressure points.

These points on pipes can be considered as "fairly fixed pressure" points	These points on pipes cannot be considered as "fairly fixed pressure" points
<ul style="list-style-type: none"> Flow point near storage tanks Flow point near a vessel with a constant level (level controlled) Flow point near a vessel with a constant pressure (pressure controlled) 	<ul style="list-style-type: none"> Flow point on a network with multiple users Flow point near a container with limited surge capacity Flow point far from fixed pressure points

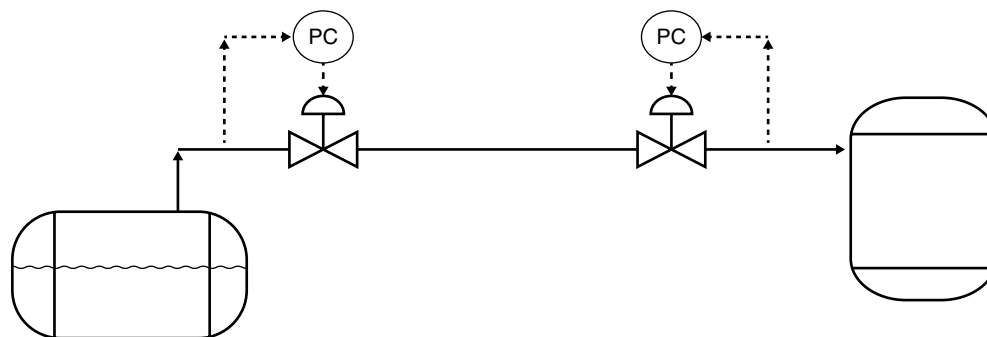


Figure 15.18 "Fixing" flow rate.

15.6.2 Controlling Multiple Pipes

In this section we talk about the control of multiple connected pipes. These arrangements can be classified into two main arrangements of flow merging and flow splitting.

15.6.2.1 Flow Merging

Stream merging could be for one of the following purposes:

- For gathering
- To provide a carrier
- For blending
- For make-up or back-up purposes (can be referred to as the “preferred source” concept).

“Gathering” is the operation where we “unify” several pipes to send the fluid to a single destination. Generally, in these cases, all the streams are the same fluid, or are compatible with each other. In this application, we are not looking for a specific flow rate on any stream; we only want to merge them together.

Stream merging for the purpose of providing a carrier is when we add one stream to another to make it easier to transfer, or to make it easier for some other final purpose. For example, adding water to lime in the powder form and making lime slurry in the majority of cases is done simply to convert the powder, which is hard to move, into a liquid, which is easy to move.

In such applications, generally only the flow rate of one stream is important, and only one stream is controlled. In the lime slurry example, we need to add enough water to the lime powder slurry to make it moveable; however, if the plan is to send lime to a reactor for a reaction, then the concentration of the lime needs to be much lower than the adequate lime concentration for conveying. As the reaction happens mainly in the liquid phase (rather than the solid phase), the lime needs to be in the liquid phase, and with diluting it we will have a better reaction with less unconsumed lime leaving the reactor. The “flow” of lime powder may be controlled upstream of its merging point, since it is a reactant in the downstream reactor.

In blending operations, several streams are merged together to achieve specific properties in the final mixed product. For example, in the final stages of refineries, they mix different proportions of light and heavy products with each other to make gasoline with specific properties, kerosene with specific properties and so on.

In factories that make orange juice from concentrate, at the start of the process, they make concentrates from oranges (by evaporation), and then they add enough water to make orange juice with a specific brix number (sweetness) and concentration. These are all examples of

blending. In blending, the flow rates of all streams are important.

In make-up or back-up applications, one (or more) stream(s) work(s) to compensate for the shortage or lack of other stream(s). Generally, all streams have a unique composition in make-up or back-up applications.

The best way to illustrate the different ways of controlling flow merging is to look at examples.

Figure 15.19 shows a stream-merging example where only one stream is flow controlled. In this example, stream A could be the fluid of importance and stream B could be the carrier.

Figure 15.20 shows a stream-merging example where stream A is the solution of importance and stream B is the carrier. Since the concentration of the active ingredient in stream A is fluctuating, to make sure that we always send a specific mass of the active ingredient, we put a composition loop control on stream A.

Figure 15.21 shows an example in which the concentration of the final product is important. This could be an example of a blending operation.

The only issue with this control system could be its slow response. As we know, the majority of composition analyzers are slow. However, it could be acceptable if the process analyzer is not very slow, which could be the case for pH or conductivity analyzers.

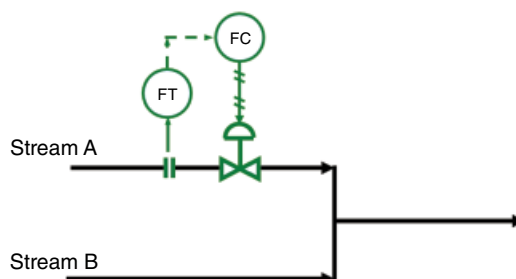


Figure 15.19 Flow merging: route of least resistance.

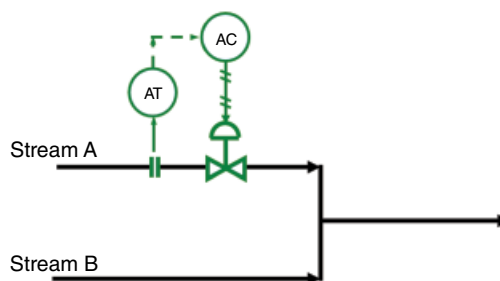


Figure 15.20 Flow merging with composition loop.

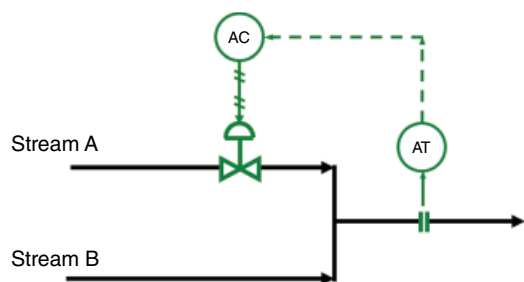


Figure 15.21 Flow merging with combined stream response.

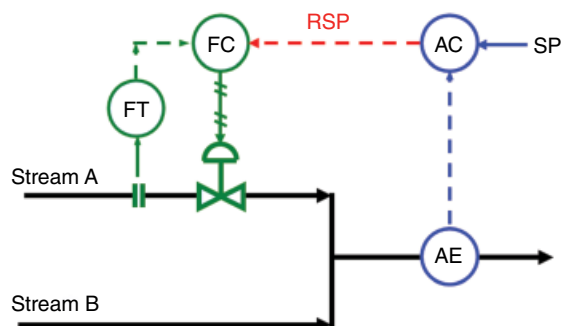


Figure 15.22 Flow merging with feedback, cascade control.

Figure 15.22 is the same arrangement as in the previous example, except that the valve on the reagent stream is controlled by feedback from the combined stream. This composition-to-flow cascade control could be beneficial, especially when stream A, the reagent, comes from a header and has fluctuating pressure and flow.

Figure 15.23 shows a blending operation in which the flow rates of both streams are important and need to be controlled. Therefore, ratio-control architecture is used. Ratio-control architecture can be shown by an FY operator, or by FFC. As a reminder, FFC stands for flow fraction control, which is the same as ratio control.

In Figure 15.24, we show the same control system as in Figure 15.23, with different symbology.

The control systems shown in Figures 15.23 (and 15.24) are ratio-control systems. As we know, ratio control is a type of feedforward system; no feedback comes from the product. In pure ratio control, there is no way to check if the control that was performed was successful or not.

Here, in Figure 15.25, we have added feedback control to the control system in Figure 15.23, in order to achieve better control. We can use the control system in Figure 15.25 if the composition of final product is very important.

The last example of stream merging is shown in Figure 15.26. In this example, a stream with a specific

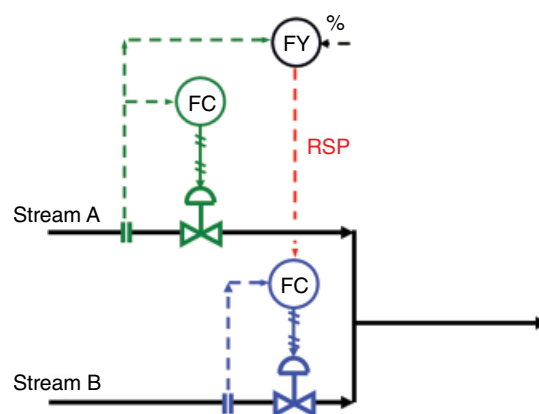


Figure 15.23 Flow merging with ratio control.

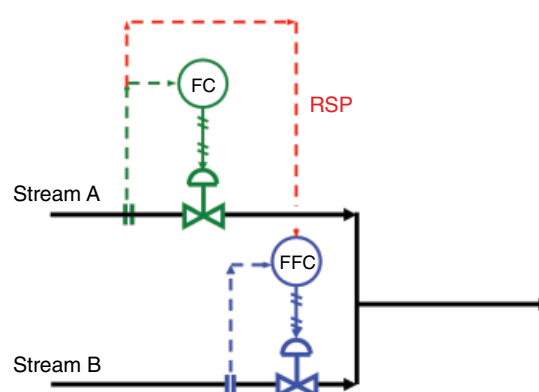


Figure 15.24 Flow merging with ratio control, other schematic.

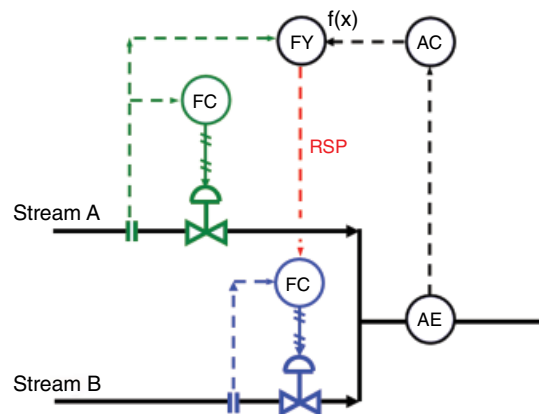


Figure 15.25 Flow merging with ratio control and composition feedback.

flow rate and a specific composition needs to be prepared. Stream A could be the reagent and stream B is most likely the carrier.

Cross-limiting control is a type of control for two merging streams when they have to be tightly attached to each other from a flow rate viewpoint. It means they should follow each other (from a flow rate viewpoint) very tightly. The flow rates of two streams could be decided to be same or follow each other with a specific ratio or specific relation.

Even though the concept of cross flow control is general and can be applied for any two merging streams, it is commonly (and arguably uniquely) used for two streams of fuel and air merging together in burner in fired heaters.

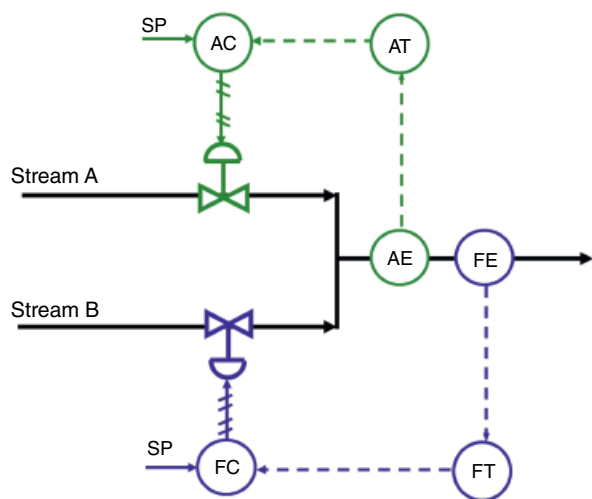


Figure 15.26 Flow merging: specific flow rate and specific composition.

Figure 15.27 shows this type of control. The building of this can be started from the schematic in the left-hand side which is a temperature-to-flow cascade control to two slave controllers. The cross-limiting control can be applied by “tying together” these two slave control loops through two override controls, which are triggered against each other, shown on the right-hand side.

15.6.2.2 Flow Splitting

In many cases, flow-splitting control is very similar to flow-merging control, just the other way around.

The main purpose of stream splitting is to send the flow to two (or more) different destinations. One famous example of stream splitting is splitting flow to go to similar pieces of equipment in parallel. Sometimes when we are distributing flow among similar parallel units, we don’t use any control systems and rely only on symmetrical piping to distribute the flow evenly.

However, this is not always the case. In particular, in cases where the receiving units are prone to plugging, like heat exchangers, or whose intention is plugging, like filters, we need some sort of control and relying on symmetrical piping is not wise.

It is not rare to see an outlet stream from a container needing to be split. In such cases, we can use the level of the container in our control loop design.

Let’s have a look at some examples that are specific to flow-splitting control.

Figure 15.28 shows the first thought that may come to mind for stream-splitting control: two branches that are independently controlled. It is rare to control a stream-splitting scenario in this way. The reason for this is that with this type of control the flow in the main header (before splitting) is fluctuating a lot. This is possibly not the concept of splitting a flow.

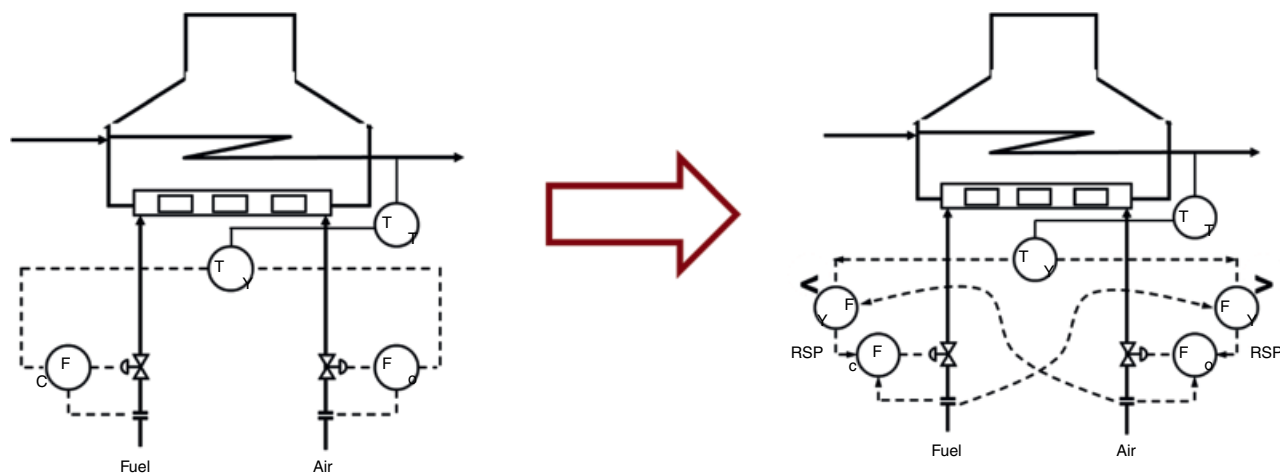


Figure 15.27 Cross-limiting control.

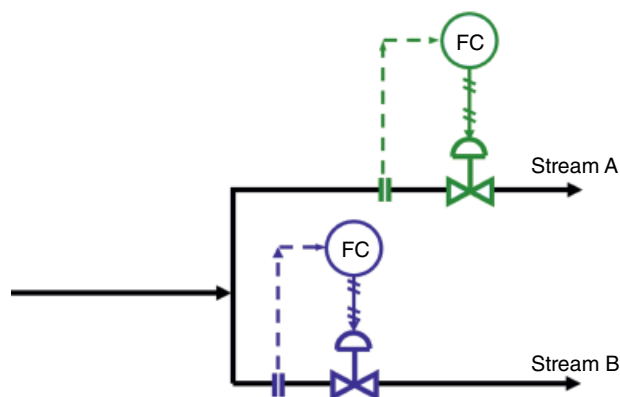


Figure 15.28 Flow splitting: independent branch control.

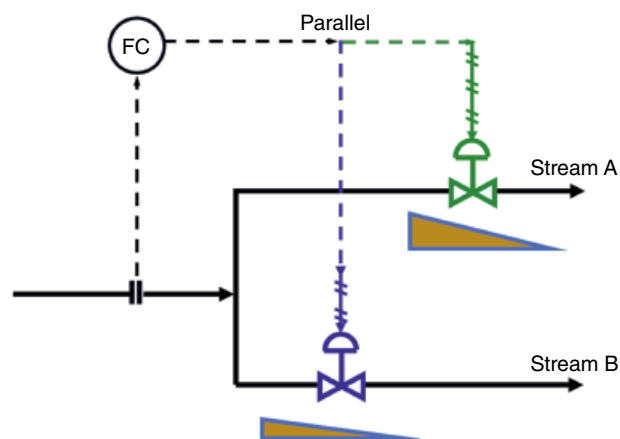


Figure 15.29 Flow splitting with parallel control.

By doing this type of control, flow surges created upstream of this control should be taken care of by – for example – placing a surge tank upstream.

Better stream splitting can be done using a parallel-range control strategy. This is shown in Figure 15.29. In this arrangement, more flow goes through stream A than stream B. The flow loop can be replaced with a level loop on the upstream container. If you choose to split the stream based on the concept of “preferred destination,” you can use the same schematic as in Figure 15.29, but using “split-range control” rather than “parallel-range control.”

The next example is a very important example of stream splitting, which is used commonly. In this arrangement, one stream is with level loop control from the upstream container, and the other stream is with a flow loop (Figure 15.30).

The most important point here is that the flow control should be on the branch with the smaller flow and never the other way around. This is because it makes the control far more accurate.

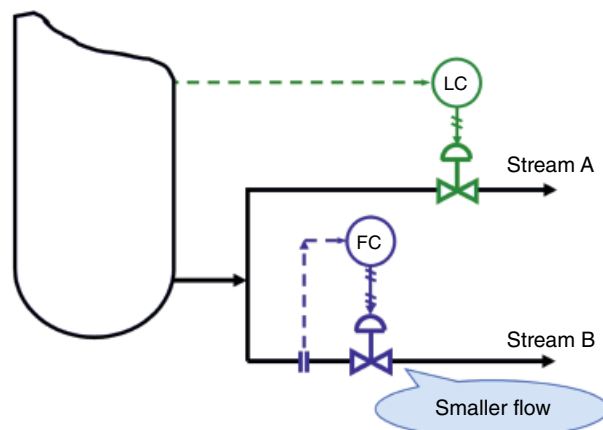


Figure 15.30 Flow splitting from a tank.

One famous example of this technique is on the top of distillation towers. In Figure 15.31, you can see two arrangements for splitting condensate from the bottom of a condensate drum.

In a typical distillation tower, the vapor coming off the top goes through a condenser and then to a drum. From the drum, the liquid is split into a reflux stream and another stream that goes to downstream equipment. The flow in each split stream will depend on the reflux ratio desired. The arrangements for either high or low reflux ratio must conform to the rule that flow control should be on the stream with the smaller flow, and level control should be on the other.

In Figure 15.31, the higher flow rate stream is shown with thicker lines for clarification.

Figure 15.32 shows flow splitting of the stream based on the “preferred destination” concept.

In this arrangement, the flow preferably goes to stream A. When there is a problem downstream that would make it difficult for stream A to take the flow, this manifests itself as an increase in pressure upstream. The higher pressure is picked up by the pressure loop on stream B, and in response the control valve on stream B starts to open to allow some flow to go through to stream B. In some designs, the control valve on stream B is replaced with a switching valve (on/off action).

15.7 Fluid Mover Control System

There are two main classes of fluid movers: liquid movers and gas movers.

The common control component for all fluid movers is capacity control. Capacity control means controlling the flow rate generated by the fluid mover. Fluid movers may have more control components in addition to capacity control.

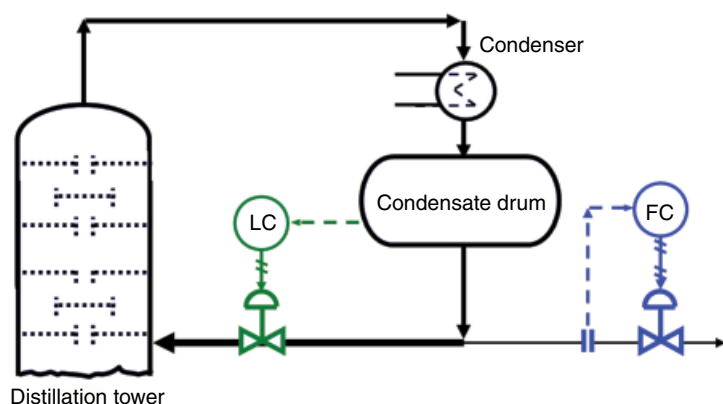


Figure 15.31 Flow splitting in a distillation column.

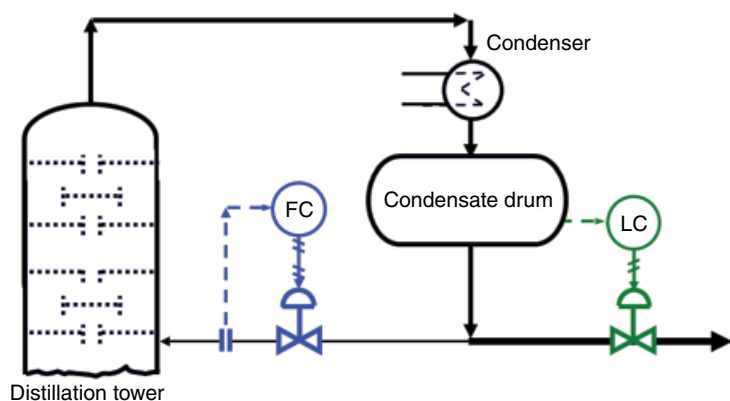


Figure 15.32 Flow splitting on flow/pressure control.

15.7.1 Pump Control Systems

There are two main classes of pumps, and the control strategies for each type are different.

15.7.1.1 Centrifugal Pump Control

When talking about pumps, we need to deal with two different, separate and independent control systems:

- Capacity control: either by control valve or variable speed drive (VSD).
- Minimum flow control. Minimum flow is the main weakness of a centrifugal pump.

15.7.1.1.1 Capacity Control


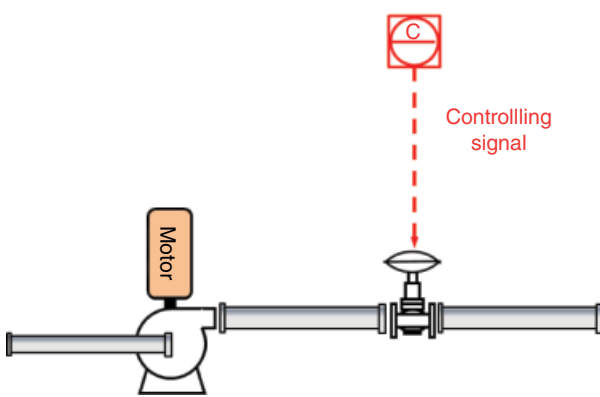

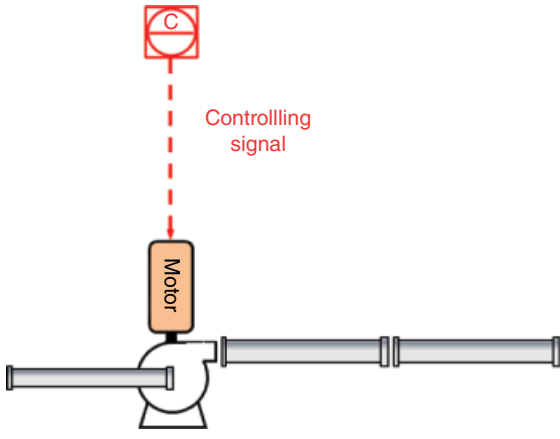
The main function of pumps is to generate flow, so flow (or capacity) needs to be controlled.

There are mainly two types of devices that can be used to adjust the flow rate: control valves and VSDs.

Control valves are generally installed on the discharge line of a pump, and VSDs are installed on the electromotor of the pump. These two methods of controlling the flow rate are very similar to two methods of controlling the speed of car we are driving. This analogy can be seen in Table 15.8.

Figure 15.33 shows the options we have for the control of a centrifugal pump. The schematics show a dashed-line loop for minimum flow control, but we will discuss this later. Option A shows capacity control using a control valve on a flow loop. Option B shows control using a signal from the flow controller to a VSD.

Table 15.8 Options for centrifugal pump capacity control analogy.

Controlling the speed of a car	Controlling flow rate
 <p>Pushing the brake pedal whenever you want to decelerate while always pushing the gas pedal!</p>	 <p>Pinching back the valve whenever you want to decrease the flow rate (without touching the pump's motor)</p>
 <p>Releasing the gas pedal whenever you want to decelerate</p>	 <p>Decreasing the flow rate of the pump (by adjusting its motor RPM) whenever you want to decrease the flow rate</p>

In option A, we basically adjust the pump's flow using a control valve on the discharge piping of the pump (discharge throttling).

In option B, we adjust the pump's flow by adjusting the rotation speed of the shaft.

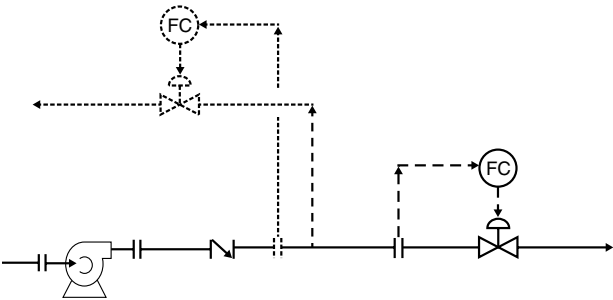
These are two completely different methods of pump control. By using a VSD, we are changing the location of a pump curve, while by using the control valve we are changing the system curve. In the majority of cases, we use a VFD (variable frequency drive) as our VSD of choice for rotary machines in plants.

Some people use the analogy of controlling the speed of a car for pump capacity control. We can see this in

Figure 15.33. Option A for controlling a car's speed is like pushing the gas pedal to the limit and pushing the brake pedal at the same time whenever we want to reduce speed. Option B for controlling a car's speed is like adjusting the speed by adjusting the position of the gas pedal.

Whenever there is more than one control method for controlling a piece of equipment, both of them can be implemented through a "split-range" control loop. For example, the capacity of a pump can be controlled by a split range between a VFD and a control valve.

Option A



Option B

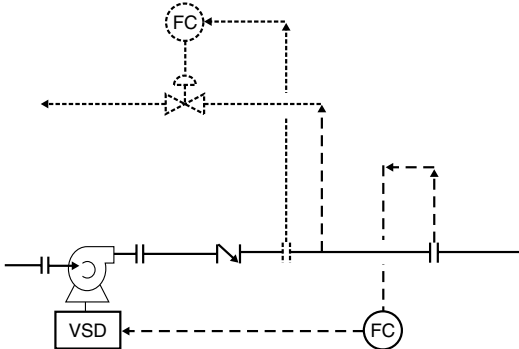


Figure 15.33 Options for centrifugal pump capacity control.

Control Valve Versus VFD

We learned that there are at least two methods of controlling pumps: by control valve and by VFD. Now the question is, which one should be used? Sometimes when we are not sure, we use both of them in the form of split-range control. If we cannot afford to use both of them (in the form of split control), we need to choose one of them. Traditionally, we use a control valve because it is an older technology; however, there are cases where a VFD works better.

The main consciously changing parameter in a water treatment plant is the flow rate. The flow rate is changing and we have to find “something” to adjust the flow rate of different pieces of equipment.

There are mainly two types of device can be used to adjust the flow rate: control valve and VSD.

A control valve is generally installed on the discharge line of a pump (or compressor) and the VSD is installed on the electromotor of the pump (or compressor).

These two methods of adjusting flow rates are named “final control elements.”

These two methods of controlling the flow rate are very similar to two methods of controlling the speed of car we are driving. This analogy can be seen below.

Table 15.9 summarizes some process reasons for using a control valve or a VFD.

Table 15.9 Options for centrifugal pump capacity control.

Control valve	VFD
<ul style="list-style-type: none"> Generates more shear on the stream. Not good for shear-sensitive liquids like oily waters, biomaterial, water carrying flows etc. Works for all types of piping circuits 	<ul style="list-style-type: none"> Generates less shear on the stream Doesn't work in systems where the majority of the pump head is used to overcome static pressure rather than pipe pressure loss

15.7.1.1.2 Minimum Flow Control

The concept of minimum flow control is shown in Figure 15.34. Let’s assume that this pump has a capacity of 200 m³ h^{−1}. The vendor specified that the minimum flow rate of the pump is 100 m³ h^{−1}. We can provide a recirculation line from the outlet of the pump back into the inlet line.

If the flow rate into the pump is 165 m³ h^{−1}, the pump is happy, since its flow is higher than the minimum flow. In this case, the sensor on the pump outlet sends a signal to the control valve on the recirculation line and it remains closed. However, if the flow drops below 100 m³ h^{−1}, for example to 80 m³ h^{−1}, the flow sensor on the outlet will send a signal to the controller to say: “I am short of my minimum required flow by 20 m³ h^{−1} and I am worried about the pump. Please open the valve enough to recirculate 20 m³ h^{−1}, so we can fool the pump into thinking that the flow is 100 m³ h^{−1}, and make it happy,” The control valve on the recirculation line will be partially opened to provide a flow of 20 m³ h^{−1}, which is sent back to the inlet to satisfy the minimum flow condition of 100 m³ h^{−1}, and prevent damage to the pump. This is the concept of minimum flow control.

It is important to recognize that this “trick” only increases the flow rate inside the recirculation loop to a number higher than 100 m³ h^{−1} to “fool” the pump. We are not able to increase the overall flow in the whole upstream and downstream piping system; the flow in those pipes is still 80 m³ hr^{−1}.

The point here is that the sensor should be placed as close as possible to the pump and within the recirculation loop.

Now the question is whether we need a minimum flow control loop for all centrifugal pumps or not. The answer is no! We don’t need minimum flow control loop for all centrifugal pumps.

The following examples of pumps may not need a minimum flow control loop [1]:

- Small pumps of less than 5 hp; they need it, but they are inexpensive so we don’t bother to put an expensive minimum flow control loop on them.

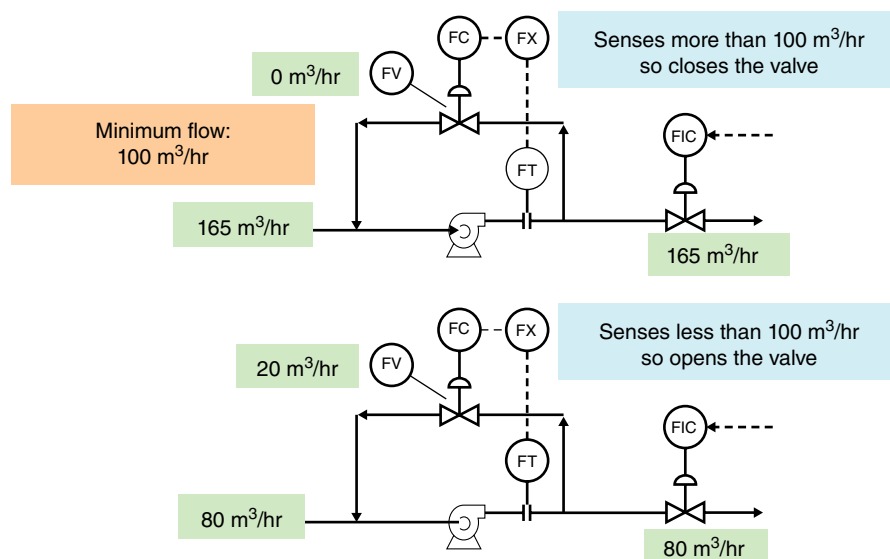


Figure 15.34 Minimum flow control.

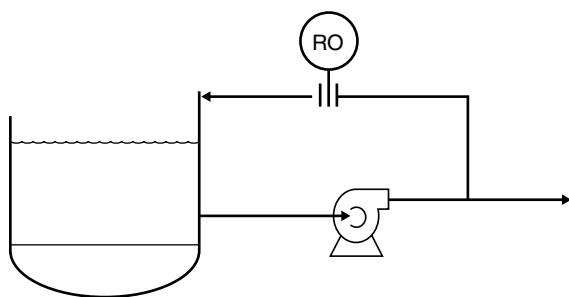


Figure 15.35 Minimum flow control by RO.

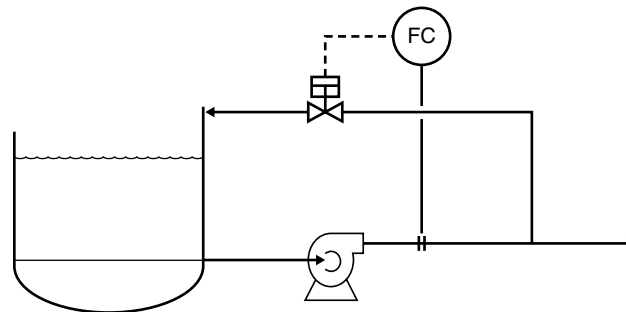


Figure 15.36 Minimum flow control by switching valve.

- Pumps in a circulating, closed system: they may not need a minimum flow control loop because the flow in such pumps is fairly constant.

However, the following examples of pumps may require minimum flow control:

- Large pumps of 5 hp or more
- Pumps on the main stream.

The rule of thumb for providing minimum flow control is as follows:

- Power < 5 hp: no minimum flow line is required, as mentioned before.
- 5 hp < Power < 10–20 hp: continuous minimum flow line. Instead of going to the expense of installing a control valve, we can put a restriction orifice (RO) on this line (Figure 15.35). By doing this, we are always recirculating a portion of flow, even in cases where flow to the pump is higher than the minimum flow and we don't really need recirculation. Thus, we are continuously wasting energy. We know that, but the pump

is so small that installing an expensive control loop for them is hard to justify. (You can picture an RO as a “frozen” control valve with a specific opening size.)

- 10–20 hp < Power < 15–30 hp: ON/OFF minimum flow line. This is a cheaper option than installing a control valve. In this case, the flow controller will just provide an ON/OFF function (Figure 15.36).
- Power > 35 hp: minimum flow line with a control valve. This is the most complicated, most expensive option, but it is the most common method of controlling minimum flow in centrifugal pumps. This type of control can be done with at least two different arrangements: with a flow loop, and with a pressure loop, as shown in Figure 15.37.

We have the option of controlling by flow sensor or by pressure sensor. Usually we go with a flow loop.

When there are two (or more) pumps in parallel and only one of them is operating and the rest are spares, a minimum spillback pipe from the common header works well.

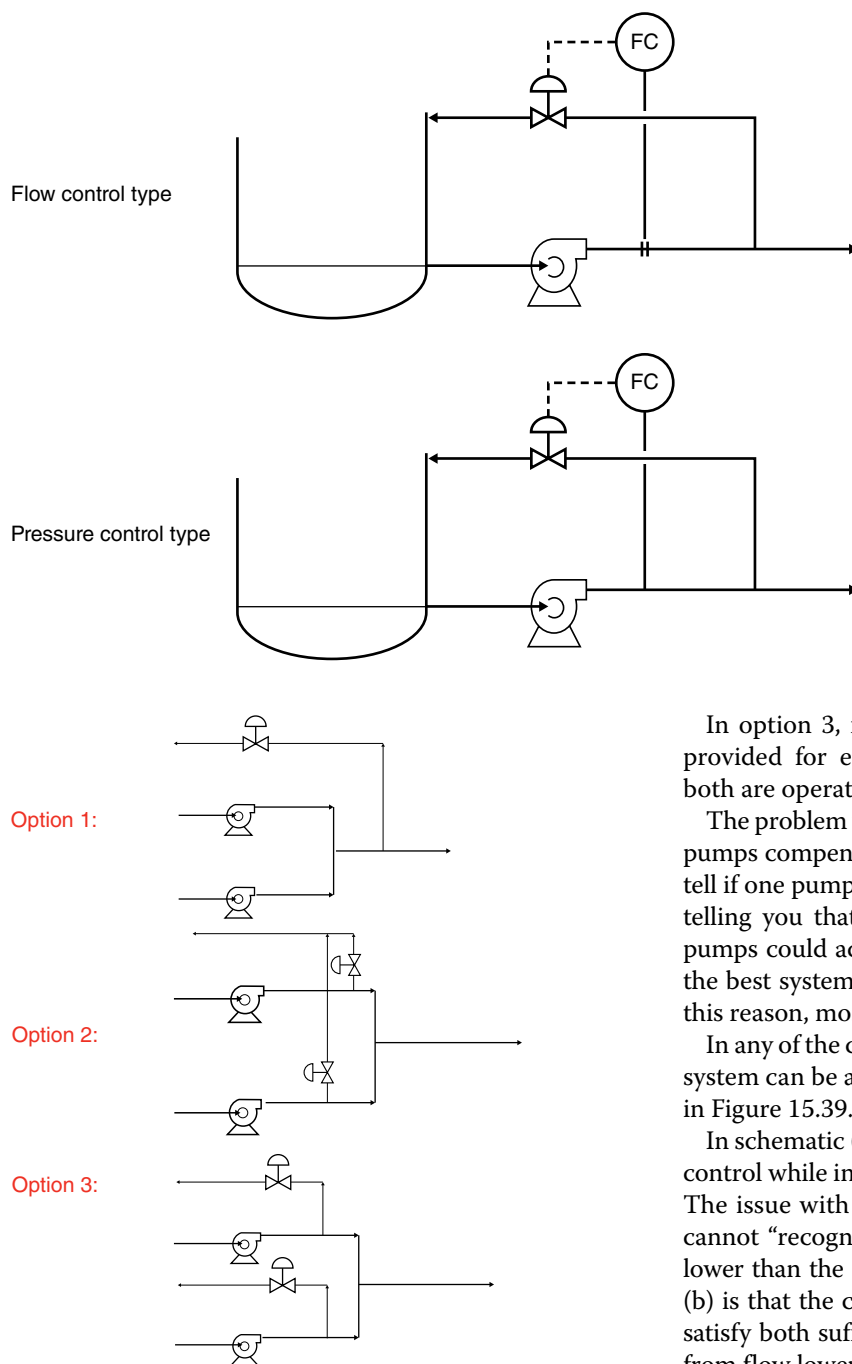


Figure 15.37 Automatic control of minimum flow.

In option 3, fully dedicated minimum flow pipes are provided for each pump in parallel arrangement and both are operating.

The problem with option 1 is that the flow from the two pumps compensate for each other and the sensor cannot tell if one pump is starving or not. So the sensor could be telling you that the system is OK, whereas one of the pumps could actually be in bad shape. Option 3 may be the best system, but it is the most expensive option. For this reason, most companies go with option 2.

In any of the cases at least two different types of control system can be applied on the minimum flow pipe shown in Figure 15.39.

In schematic (a) the total flow is used for the purpose of control while in schematic (b) the minimum flow is used. The issue with schematic (a) is that the control system cannot “recognize” the pump is suffering from the flow lower than the minimum flow. The issue with schematic (b) is that the control system cannot provide the flow to satisfy both suffering pumps if both pumps are suffering from flow lower than the minimum flow but with uneven difference to the minimum flow.

Figure 15.38 Minimum flow pipes for parallel operating pumps.

The question is what the minimum flow should be when there are parallel operating pumps. We have three options for providing minimum flow lines for parallel operating pumps, as shown in Figure 15.38.

In option 1 one, a shared minimum recirculation pipe is provided for two operating pumps.

In option 2, dedicate recirculation pipes are provided at the beginning but they are merged in downstream before getting to the upstream reservoir.

15.7.1.2 Positive Displacement (PD) Pump

Because more than 80% of pumps used in the process industry are centrifugal pumps, many people are not familiar with how to control PD pumps. You cannot use the same arrangement for controlling a PD pump as for a centrifugal pump; it does not work. If you put a flow loop on the discharge side of the pump, it won't change the flow at all.

There are two main groups of PD pumps: reciprocating and rotary. Each has its own symbol, but I have invented

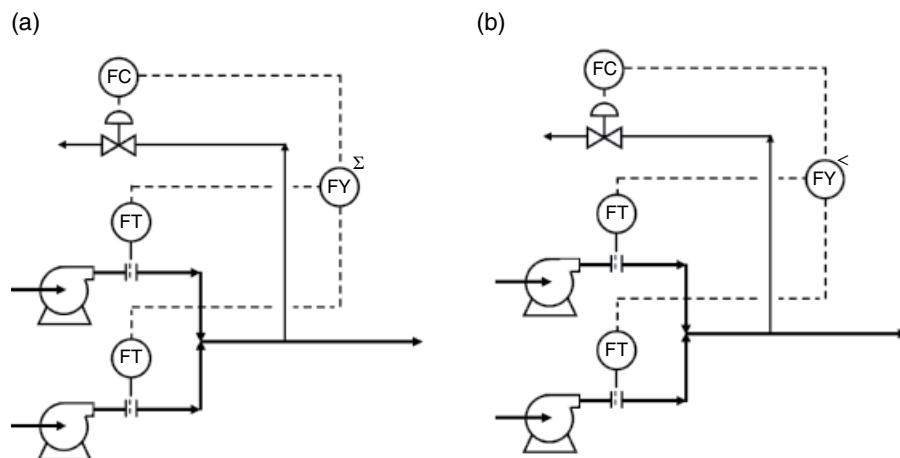


Figure 15.39 Control of minimum flow pipes for parallel operating pumps.

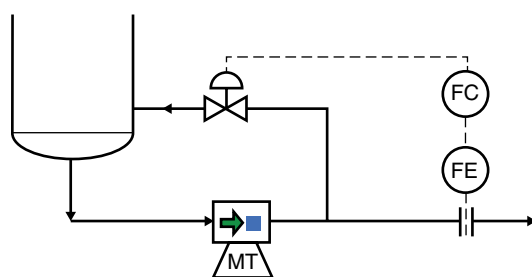


Figure 15.40 PD pump control by recirculation line.

my own symbol, shown with my initials, MT, to cover the whole range of PD pumps. This is not a symbol that you should use on a P&ID because it is not recognized.

Positive displacement pumps are not sensitive to minimum flow, so all we have to worry about is capacity control. For capacity control, again we have two options (but these options are different from the options for centrifugal pumps): by recirculation pipe (and not discharge throttling) and by VSD.

15.7.1.2.1 PD Pump Control by Recirculation Pipe

To understand the way we can control the capacity of a PD pump, you need to picture a PD pump as a “stubborn” system, i.e. you cannot “convince” it to decrease its capacity. The only thing you can do is that you divert a portion of its capacity out of the main discharge pipe.

There are at least two arrangements for capacity control of PD pumps using a recirculation pipe. The first arrangement is shown in Figure 15.40.

This arrangement shows how we can control the capacity of the PD pump by using a recirculation line. The flow sensor is located on the discharge line, outside of the recirculation loop.

If the rating of the pump is $10 \text{ m}^3 \text{ h}^{-1}$ and the equipment downstream is only drawing $8 \text{ m}^3 \text{ h}^{-1}$; this means we need to recirculate $2 \text{ m}^3 \text{ h}^{-1}$.

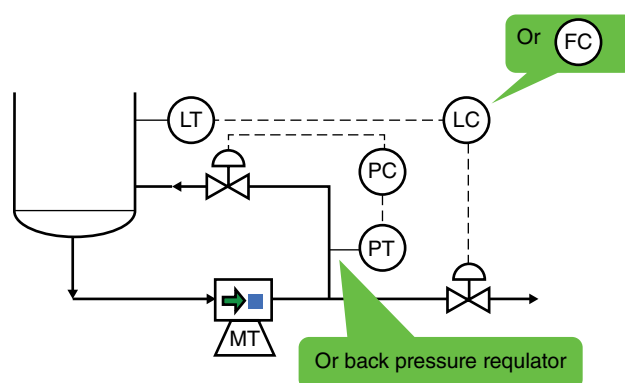


Figure 15.41 PD pump control by recirculation line.

Even though this arrangement is similar to a minimum-flow configuration, the design basis for the recirculation pipe is different from a minimum-flow protection pipe. Also, the flow sensor is located outside of the recirculation loop.

The arrangement shown in Figure 15.41 is somewhat complicated and not popular for that reason.

Here you can see a control valve on the outlet (discharge) of the PD pump, and you may be surprised that “it doesn’t work”! You are right, controlling the capacity of a PD pump with discharge throttling doesn’t work, but here we have added a recirculation pipe to make it workable.

The control loop on the discharge of a PD pump could be a level loop or a flow loop, depending on the situation. The control loop on the recirculation pipe should be a back pressure control loop. We may choose to replace the “back pressure control loop” with a “back pressure regulator.”

15.7.1.2.2 PD Pump Control by VSD

The second option for capacity control of a PD pump is using a VSD, or a VFD (Figure 15.42). Here, we know that the PD pump is stubborn, but we overcome this by managing it from its “root”; we control the speed of the pump shaft.

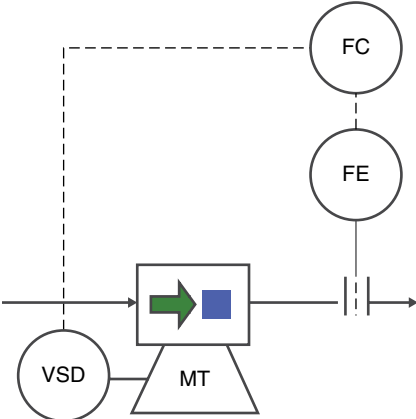


Figure 15.42 PD pump control by VSD.

The capacity of the pump is controlled by a flow loop. A flow sensor on the discharge line sends a signal to the flow controller, which, in turn, sends a deviation signal to the VSD on the electromotor of the PD pump.

15.7.1.2.3 PD Pump Control by Stroke Adjustment

A third type of PD pump control is only for reciprocating pumps. In the reciprocating type of PD pumps, a member does a forward and backward movement repeatedly in order to achieve the pumping.

The third method of controlling the capacity is applicable only to reciprocating pumps: adjusting the span of the member’s reciprocating movement, or simply adjusting the stroke.

This can be done by attaching a servomotor to the “stroke-adjustment lever” of the pump.

The signal from the flow controller can increase or decrease the stroke length, which will adjust the flow rate. This control is shown in Figure 15.43.

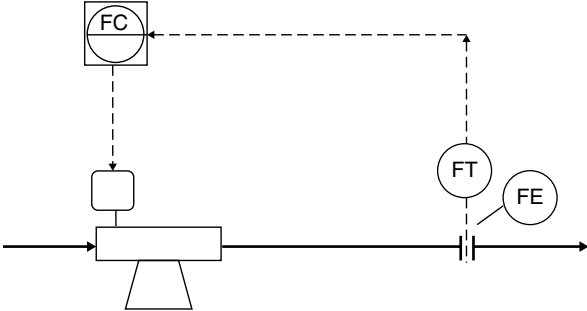


Figure 15.43 PD pump control by adjusting stroke length.

15.7.2 Gas Mover Control Systems

There are two main types of gas movers:

- Dynamic. These can be either axial or centrifugal.
- Positive displacement. These are either reciprocating or rotary.

Control systems for gas movers are very similar to pump control systems; both need to have capacity control. Whereas we don’t need minimum protection for PD pumps, we do need it for PD gas movers. Minimum protection for pumps is referred to as “minimum flow control,” and for gas movers, it is called “surge protection.”

Again, for centrifugal gas movers, there are two control systems needed: capacity control, and anti-surge control.

15.7.2.1 Capacity Control Methods for Gas Movers

There are five main types of control for gas movers. These are: recirculation pipe (or spillback pipe), speed control, suction throttling, discharge throttling, and a combination of control methods based on the structure of the gas mover.

Table 15.10 shows which methods are technically applicable for which types of gas mover.

- Recirculation. This is a general control method that can be used for any type of gas mover. In this

Table 15.10 Control methods for gas movers.

	Dynamic		Positive displacement	
	Axial	Centrifugal	Reciprocating	Rotary
Recirculation	General control method			
Speed control via VSD	General control method			
Suction throttling	✓	✓	✓	✓
Discharge throttling	✓	✓	✗	✗
Special control	Applicable	Applicable	Applicable	Applicable

method, a pipe connects the discharge side of gas mover to its suction side (spillback pipe). A control loop is placed on the spillback pipe. This arrangement looks very similar to the “minimum flow protection” system in pipes, but there are plenty of differences in the details.

- **Speed control (VSD).** Again, you can use this method of control for all gas movers. In this method, the speed of the gas mover’s shaft is adjusted through a mechanism. As was mentioned, the “mechanism” could be a VFD (for electric motors as the drive) or a governor (for turbines as the drive).
- **Suction throttling.** Suction throttling means putting the control valve on the suction side of the gas mover. With a pump, you have the danger of cavitation if you place the control valve on the suction side, but cavitation is not an issue with gas movers.
- **Discharge throttling.** This means that the control valve is placed on the discharge side of the gas mover. This method of control is acceptable to use with dynamic gas movers, but not with PD gas movers.
- **Special control method.** Unlike the previous four control methods, this is not a single control method. Depending on the type of gas mover, a few extra

control methods could be available. We call all of these “gas mover-specific” control methods “special control.” For example, in axial gas movers such as blowers and air coolers, we can adjust the flow rate by changing the pitch or angle of the blades.

In the above paragraph, I explained the different types of control that are “technically” possible; however, not all of them are economically acceptable in each application. Table 15.11 gives the applicability of each of the control methods mentioned.

The items in Table 15.11 are expanded below:

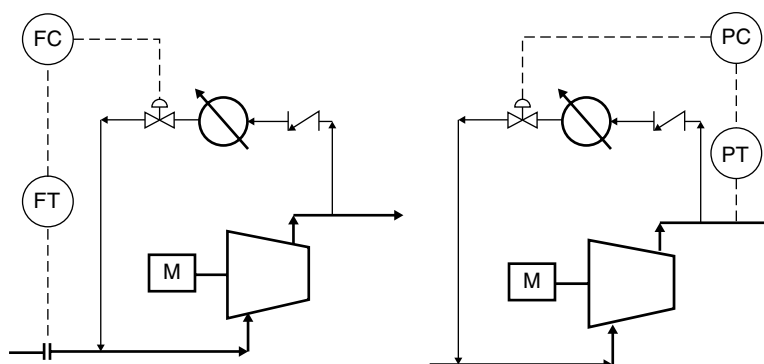
- **Recirculation.** This method is not very economical because when you recirculate back to the suction side, you are expanding the gas and therefore wasting energy (Figure 15.44).

Here we have a recirculation line with a check valve. We have included a cooler on the line because with compression, you get an increase in temperature and you may need to cool the recirculated gas before it goes back into the suction side of the gas mover. We also have a control valve on this line to prevent all of the flow from going through this attractive route of less resistance.

Table 15.11 Control methods for gas movers.

	Dynamic		Positive displacement	
	Axial	Centrifugal	Reciprocating	Rotary
Recirculation	Not economical, especially for high-pressure applications			
Speed control via VSD	General control method			
Suction throttling	Common	Common	Not good practice	Not good practice
Discharge throttling	Not economical	Not economical	✗	✗
Special control	<ul style="list-style-type: none"> • Blade pitch adjustment: most common 	<ul style="list-style-type: none"> • Guide vane adjustment 	<ul style="list-style-type: none"> • Suction valve unloading: most common • Clearance pocket adjustment 	<ul style="list-style-type: none"> • Slide valve adjustment

Figure 15.44 Recirculation control for a gas mover.



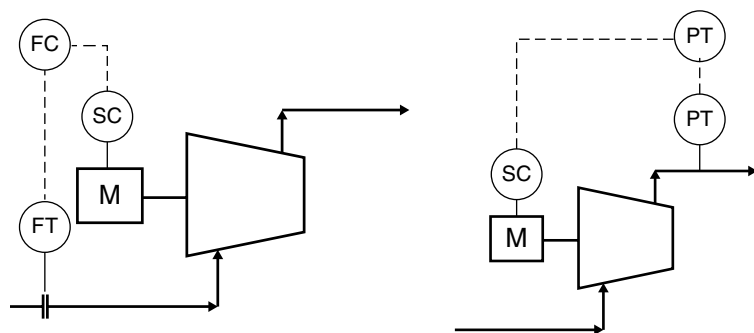


Figure 15.45 Speed control of a motor-driven gas mover.

There are two options for a control loop here: flow loop or pressure loop, and as flow and pressure are interrelated, both of them are theoretically acceptable. The decision of whether to use a flow control loop or a pressure control loop could be based on some other parameters. For example, pressure sensors generally have higher rangeability and if your flow surges a lot a pressure loop may be a better option.

The other case is when you are dealing with very high discharge pressures. In such cases, high design pressure flow meters could be overly expensive and not justifiable. This may direct you to use a pressure control loop instead of a flow control loop.

- Speed control (VSD). This method can be used with all gas movers.

Figure 15.45 shows the control of a gas mover by VSD when its drive is an electric motor.

Some companies use the symbol “SC” in their schematics, meaning “speed control.”

Some other companies use the acronym “VFD.”

Figure 15.46 shows the same VSD control on a steam turbine-driven gas mover. Here the device that works to change the speed of turbine shaft is a “governor.”

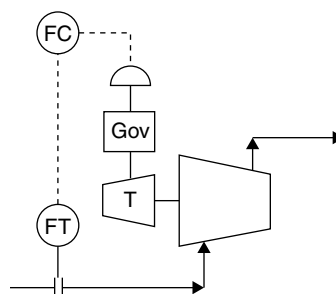


Figure 15.46 Speed control of a turbine-driven gas mover.

- Suction throttling. Suction throttling can be used for all types of gas movers but is really only popular with centrifugal gas movers (Figure 15.47).
- Discharge throttling. Although this method can be used for all dynamic type gas movers, it is not economical. By adjusting the valve on the discharge side of a gas mover, we are partially expanding a gas that we had already compressed (and spent money to do so) through the gas mover (Figure 15.48). This is a waste of energy. However, this is not much of a factor when dealing with gas movers with small compression ratios, such as fans and blowers.

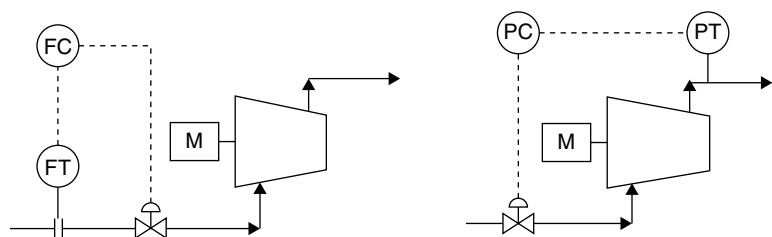


Figure 15.47 Suction-throttling control of a gas mover.

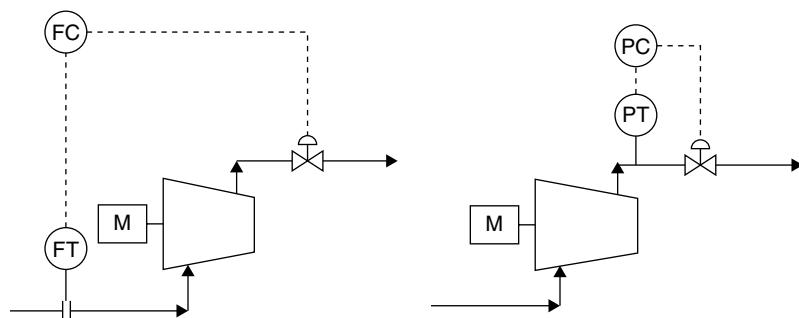


Figure 15.48 Discharge-throttling control of a gas mover.

- **Special control method.** This is a combination of capacity control methods for gas movers that uses some adjustable element in the structure of the gas mover to control its capacity. To be able to explain all of these methods, one needs to know the detailed structure of gas movers, which is beyond the scope of this book. So here, we instead explain only two simple “special control methods.”

Figure 15.49 shows “blade pitch control,” which is a special control method used only for axial gas movers. One famous application of an “axial gas mover” is in air coolers.

In the air cooler shown below, the outlet temperature of the hot stream is adjusted by the air flow rate. The air flow rate is adjusted by the angle (pitch) of the propeller blades.

In the second example, in Figure 15.50, another special control method is shown that is applicable only to

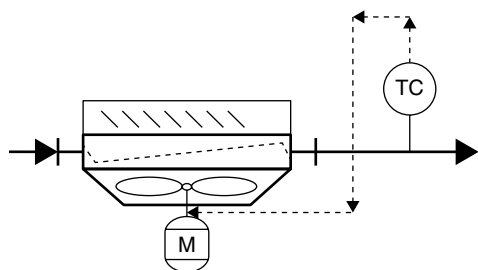


Figure 15.49 Discharge-throttling control of a gas mover.

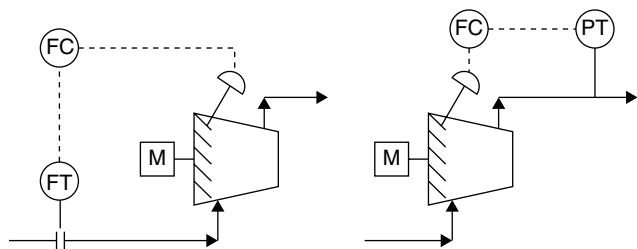


Figure 15.50 Discharge-throttling control of a gas mover.

centrifugal gas movers. In centrifugal gas movers, there is an element installed in the inlet of the gas mover, called a “guide vane.” By controlling the position of the guide vane, the capacity of the gas mover can be adjusted. To have a better picture of this, you can visualize the guide vane as being similar to a built-in control valve on the suction side of the gas mover.

15.7.3 Anti-Surge Control

When we talk about surge control, we have the same problem as with centrifugal pumps, i.e. providing protection against a minimum condition. For centrifugal pumps, this means providing minimum flow control, but for an axial or centrifugal gas mover, we have to deal with a surge problem by providing anti-surge protection control.

Figure 15.51 shows a simple anti-surge system.

It should be noted that the anti-surge control could be more complicated than what is shown in Figure 15.49. This is because surge is one of the fastest phenomena in process plants.

We don’t need an anti-surge system for low-pressure dynamic gas movers like fans or blowers. We don’t need them for PD gas movers either.

The rule of thumb to use is that anti-surge protection should be applied if the pressure differential around the dynamic gas mover is more than 15 kPa, and the gas mover’s required power is more than 150 BHP (brake horsepower).

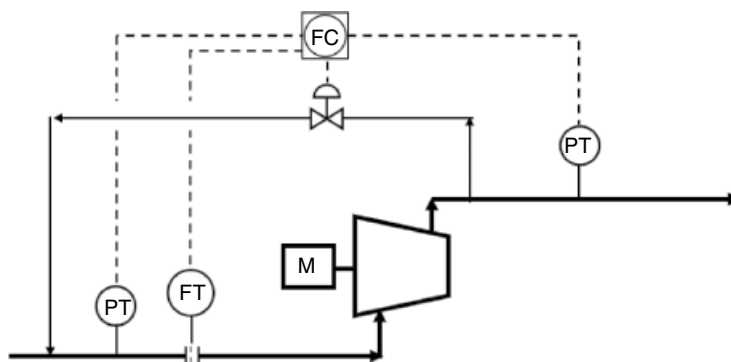
15.7.4 Lead-Lag Operation of Fluid-Movers

The lead-lag concept can be applied to other equipment, but usually it is used for fluid movers and is common for pumps.

Up to now, we have learned that when there are two pumps in parallel, there could be two cases:

- 1) One pump operating and another pump spare ($2 \times 100\%$ sparing philosophy)
- 2) Both pumps operating ($2 \times 50\%$ sparing philosophy).

Figure 15.51 Anti-surge system on a centrifugal compressor.



Here, I want to tell you that there is a third type of operation for parallel pumps: lead–lag operation. There are two things to know about lead–lag operation: (i) both pumps are not operating all the time, and (ii) it is not the case that the “spare” pumps are put into operation when the first pump fails.

Therefore, in lead–lag operation of parallel pumps, the sparing philosophy is neither $2 \times 100\%$, nor $2 \times 50\%$.

In lead–lag operation, two (or more) pumps work together – as a team – to do one specific task. At the beginning, only one pump starts to run (the “lead pump”), and when/if it cannot manage the situation, the second pump (the “lag pump”) starts to run automatically to help the first pump.

The other type of lead–lag operation is the type where the lead pump is operating continuously and the lag pump comes on to the scene when the lead pump cannot keep up with the flow.

A lead–lag pump system is normally used for vessels that have an unpredictable level because of unpredictable flow fluctuations, such as in a sump.

This is a two-pump system, where the first pump, or “lead” pump, may be working continuously. When the liquid in the sump rises to a certain level, the second, or “lag,” pump kicks in. When the level in the sump drops, the pumps will shut down in reverse order.

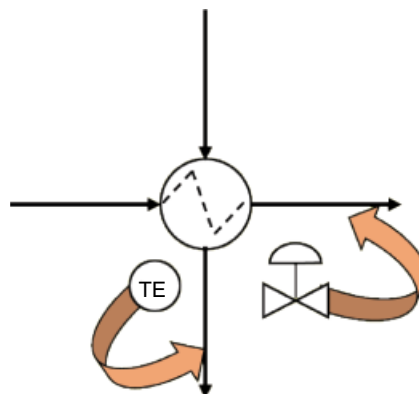


Figure 15.52 Heat exchanger control.

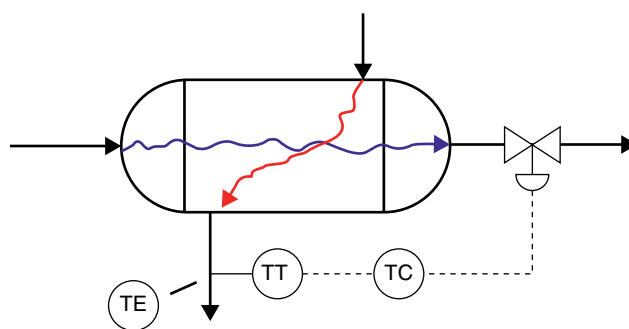


Figure 15.53 Direct control of a heat exchanger.

15.8 Heat Transfer Equipment Control

In this section we cover the control of heat exchangers and fired heaters.

15.8.1 Heat Exchanger Control System

Since the main duty of a heat exchanger is to adjust the temperature of a stream (the “target stream”), the temperature of the target stream needs to be controlled. This can be achieved by placing a temperature sensor on the target stream at the outlet of the heat exchanger.

What would be the location of the control valve? Generally speaking, the location of the control valve is on the other stream, i.e. the non-target stream. Why is this? Well, if we have a target stream whose temperature is important to us, then its flow rate is most likely important to us too, so it doesn’t make sense to put a control valve on the target stream (Figure 15.52).

This is a very important point in controlling heat exchangers.

There are two main methods of controlling heat exchangers:

- 1) Direct control: by throttling the non-target stream
- 2) Bypass control: by diverting a portion of a non-target stream to bypass the heat exchanger.

Although the first solution is easier and cheaper, there are some cases where we have to deviate from method 1 (direct control), and use method 2 (bypass control).

15.8.1.1 Direct Control System

The simplest type of control for heat exchangers is direct control (Figure 15.53). This type of control is the best for utility heat exchangers but they are not very common for process heat exchangers.

In this schematic, we have one hot stream and one cold stream. We have a temperature sensor on the discharge line of the hot stream because this is the temperature that we want to control. The temperature loop controls a control valve on the discharge side of the cold stream.

When designing heat exchanger control loops, we generally follow these rules:

- Sensor. The temperature sensor should be located on the target stream (outlet side).
- Control valve. This is usually located on the other stream. It is also placed on the outlet rather than on the inlet of the stream.

With this simple arrangement, we can control the heat exchanger.

However, let’s see the features of this control system: to be able to control the red stream, we throttle the

horizontal cold stream. By throttling the horizontal cold stream, we are changing its flow rate and its temperature. You may ask, “I can see that we are changing the flow rate of horizontal cold stream, but how can we change its temperature?”

The answer is that when you change the flow rate of the horizontal cold stream, you are in essence changing its residence time in the heat exchanger. Thus, the horizontal cold flow would end up having either more or less contact with the red stream, so its temperature will change too.

So, this control is acceptable as long as the horizontal cold stream is not an important stream and doesn't have another responsibility downstream of this heat exchanger. This is the case for utility heat exchangers. Basically, direct control of heat exchangers can be done only on utility streams. If we implemented this type of control around a “process” heat exchanger and on a process stream, it would be very troublesome. The equipment downstream of the throttled process stream would likely not be very happy about an ever-changing flow and the temperature of the process flow comes to the downstream unit.

This is not an ideal control system for a process heat exchanger because a fluctuating flow rate on the process stream may negatively affect the operation of downstream equipment.

To summarize, direct control is fine for utility heat exchangers, but it is not ideal for process heat exchangers.

One example of direct control is shown in Figure 15.54. In this example, cascade control is used.

Temperature-to-flow cascade control could be used if the response time of a simple loop is long and not acceptable.

15.8.1.2 Bypass Control System

To solve the problem mentioned above, a control system for process heat exchangers has been developed and is shown in Figure 15.55.

Figure 15.53 shows the same setup for heat transfer between a hot stream and a cold stream, with the target temperature on the hot stream. However, in this arrangement a portion of the cold stream is diverted to bypass the heat exchanger.

We still have a temperature sensor on the outlet line of the hot stream, but the temperature loop controls a valve on the bypass line and not on the outlet line of the cold stream. Here, we still control the temperature of the target stream by throttling the non-target stream. However, by putting the control valve on the bypass stream rather than on the main stream, we effectively eliminate the flow fluctuation of the non-target stream; whenever the control loop asks for a less cold stream (horizontal cold stream), instead of decreasing the horizontal cold stream (like what we did in the direct-control system), we are bypassing a portion of flow around the heat exchanger. By using this trick, we satisfy the orders of the control loop and, at the same time, we haven't changed the total flow rate of the horizontal cold stream.

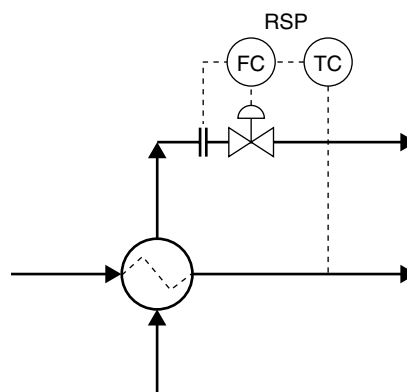


Figure 15.54 Heat exchanger direct control – cascade control example.

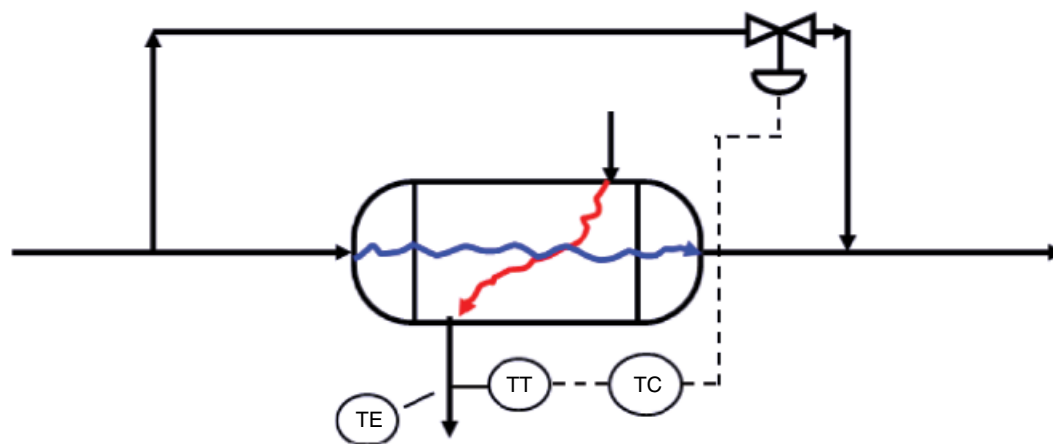


Figure 15.55 Heat exchanger bypass control.

This is a better arrangement when the horizontal cold stream is a process stream and has another responsibility downstream of this heat exchanger.

However, the temperature of the horizontal cold stream is still changing and, if it is not tolerable, we cannot do anything for it in this heat exchanger.

When the fluctuation in temperature may be too great for the equipment downstream, we can solve this by putting another heat exchanger in series with this one to control the temperature for downstream equipment.

The second heat exchanger must be a utility heat exchanger.

We can have another variation of the arrangement in Figure 15.56, by putting control valves on both the discharge and bypass streams, as shown below.

We have already shown split-range control of valves. In order to maintain a constant flow rate when we have flow from both streams at the same time, we need to go with parallel control.

We could substitute the two control valves in Figure 15.56 with a single three-way valve, as shown in Figure 15.57.

The disadvantage of using three-way valves is that they are not available in large sizes. They are also not suitable when they receive a stream with large temperature variations. This will cause different rates of expansion inside the valve, which inevitably leads to leakage.

One very interesting variation of bypass control is shown in Figure 15.58.

In the arrangement shown in Figure 15.55, we have a pressure differential controller (PDC) on the bypass line. If we have a situation where the pressure drop across the heat exchanger is large, most of the inlet flow will choose to go through the bypass line, where the pressure drop will be smaller. To avoid this, we can put in a PDC to control the pressure drop on the bypass line and thereby control the flow rate through that line.

The situation where it is a big advantage to use a PDC on a bypass line is where you have a number of heat exchangers operating in parallel. If you didn't use a PDC, you would have to have a bypass line with a control valve for each and every heat exchanger. However, if you use a PDC, you only need to install one bypass line across the whole bank of parallel heat exchangers, with one control valve.

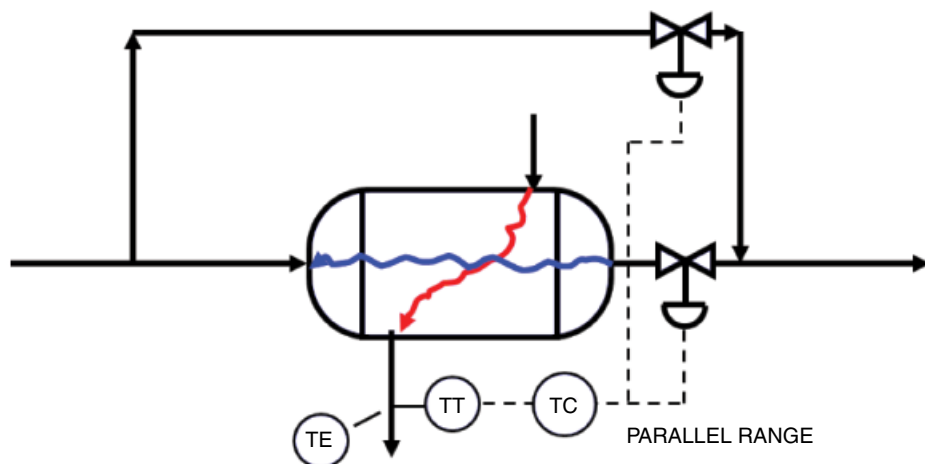


Figure 15.56 Heat exchanger bypass control with two control valves.

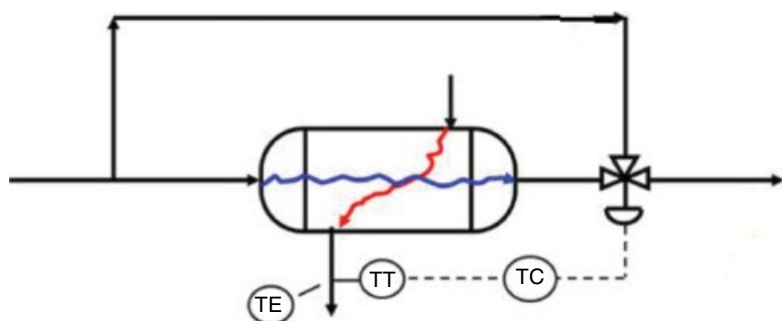


Figure 15.57 Heat exchanger bypass control with a three-way valve.

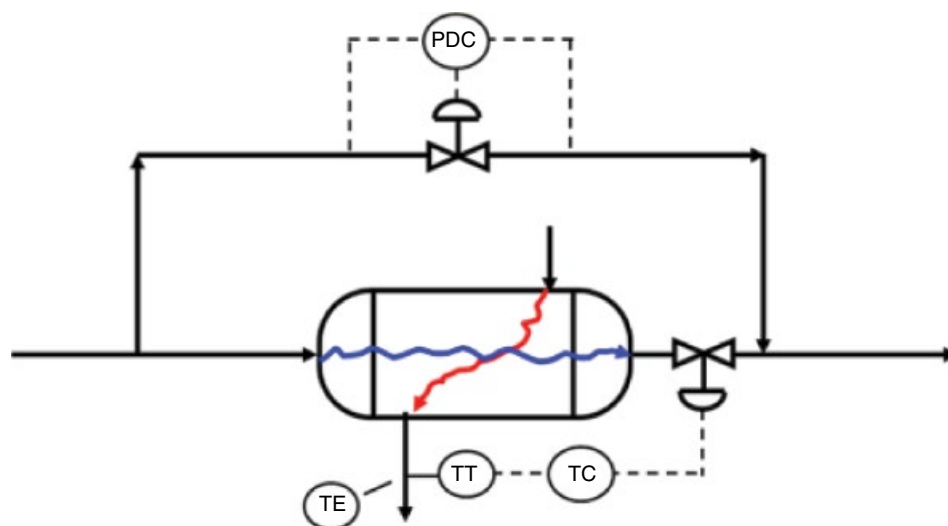


Figure 15.58 Heat exchanger bypass control with a PDC.

Temperature Control of Reactors

Heat transfer doesn't necessarily happen only in heat exchangers and then temperature control. In reactors, temperature control may be important if the reaction is largely exothermic (heat-emitting) or endothermic (heat-absorbing).

In the below schematic examples of reactor temperature control are shown. In some of the examples below, the temperature of the reactor is controlled using a jacket around the reactor, while in the other examples, heating and/or cooling coil(s) are used inside the reactor.

In this example, we have a reactor with an exothermic reaction. The reactor has a jacket for cooling water, Figure 15.59.

We want to control the reactor temperature, so we have a temperature sensor on the product line, which is connected to a loop to control a valve on the cooling water inlet stream.

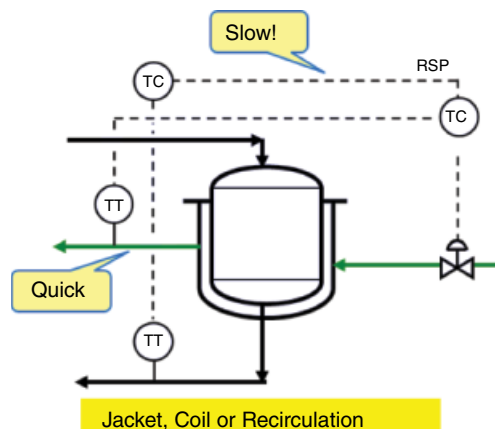


Figure 15.59 Reactor control – jacketed reactor.

However, as we know, the response time for a temperature loop is slow, especially in this case, since it involves the temperature of a bulk fluid. This arrangement could be acceptable if the speed of the loop is not an issue, for example if the reaction is very slow. In the majority of practical cases, the reaction speed (reaction kinetics) is so high that such a simple arrangement wouldn't provide good control. We can speed up the response time by using another, faster temperature loop to act as a slave for this loop.

The temperature loop of jacket water is faster because the volume of water in the jacket is much lower than the fluid in the reactor (a layer of water versus a bulk fluid). In this improved arrangement, the temperature loop on the product line acts as a master to provide a RSP for the temperature controller of the slave loop. In effect, we have a temperature-to-temperature cascade control system.

Instead of having a jacketed reactor, we can replace this arrangement with the one shown in the bottom right of Figure 15.60. To control the temperature, we divert a

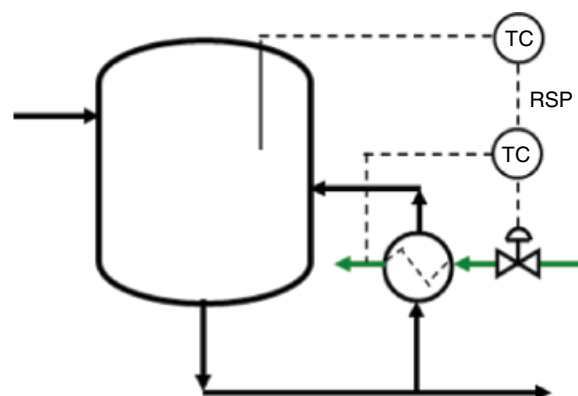


Figure 15.60 Reactor control – external heat exchanger.

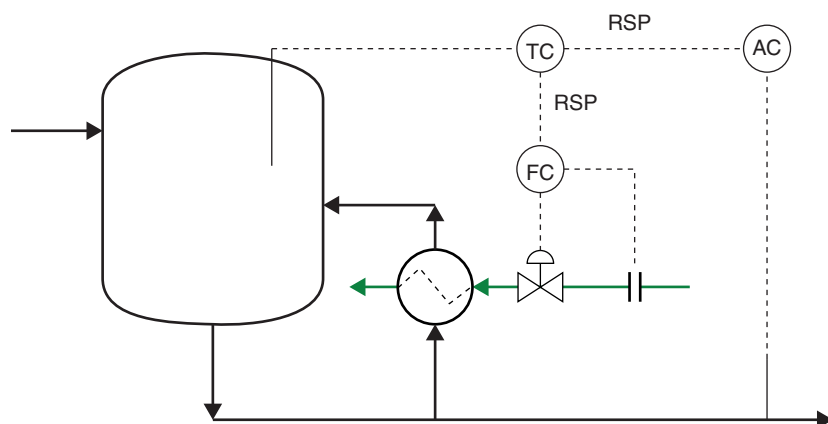


Figure 15.61 Reactor with three level cascade control.

portion of the product flow through a cooler and send it back to the reactor. The control mechanism is still the same cascade system, but instead of a master loop on the product line, we have a loop from a temperature sensor on the reactor itself.

Up to now, we have discussed the heat exchanger aspect of a reactor; we control the temperature of the reactor to get the best reaction yield. Yield means the ratio of product to feed on a mass basis.

However, we can move to a more complicated control system to “attach” the temperature loop to the real purpose of the reactor: the product quantity and quality.

Figure 15.61 shows an example of three-level cascade control, specifically composition-to-temperature-to-flow control.

It’s perfectly acceptable to control a reactor by temperature alone because temperature has a direct influence on composition. However, where composition is a critical parameter, it may be necessary to check the composition, or purity, of the product stream. So, we install a process analyzer with a control loop to provide an RSP for the temperature controller, which in turn provides an RSP for the flow controller to adjust the valve on the utility line to the cooler. This example can be repeated with replacing recirculating stream with steam coil or steam jacket.

Now let’s look at a more complicated reaction: a reaction that could be exothermic up to a certain point, and then becomes endothermic. In such a case, to control the temperature, we need both a cold stream and a hot stream.

The schematic in Figure 15.62 is almost the same as that in Figure 15.59, but it shows the provision of both hot and cold streams for the utility line to the jacketed reactor. These streams are controlled by split-range control; if the reaction is exothermic we would set the cold stream to operate at 0–50% and the hot stream to operate from 50 to 100%. Conversely, if the reaction is predominantly endothermic, the split-range control

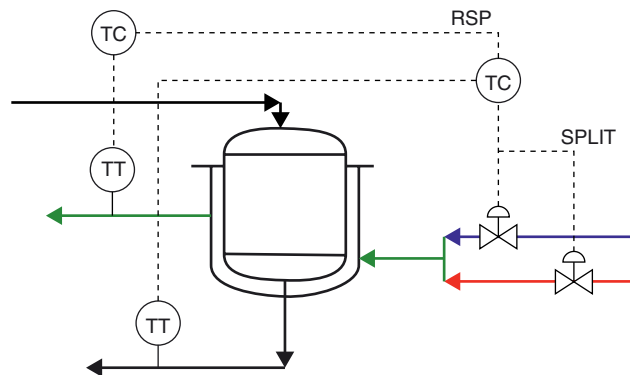


Figure 15.62 Reactor control with split-range control on utility line.

would be set the other way around. Again this example can be repeated with replacing recirculating stream with steam coil or steam jacket.

15.8.1.3 Control of Heat Exchangers Experiencing Phase Change

We have discussed two methods of heat exchanger control: direct control and bypass control. Although for heat exchangers involving a phase change, i.e. liquid to gas or gas to liquid, we can apply either of these methods of control but putting sensor or control valve on the phase-changing stream is not as simple as non-phasing stream. However, the good news is that a third way is available for phase-changing heat exchangers. On the one hand, we love to use direct or bypass control on a phase-changing heat exchanger, but on the other hand, it has complications and at the same time there is another opportunity available for phase-changing heat exchangers, which is “partial flooding control.”

The discussion below involves the specific example of a steam heater; however, it is applicable for all different types of phase-changing heat exchangers.

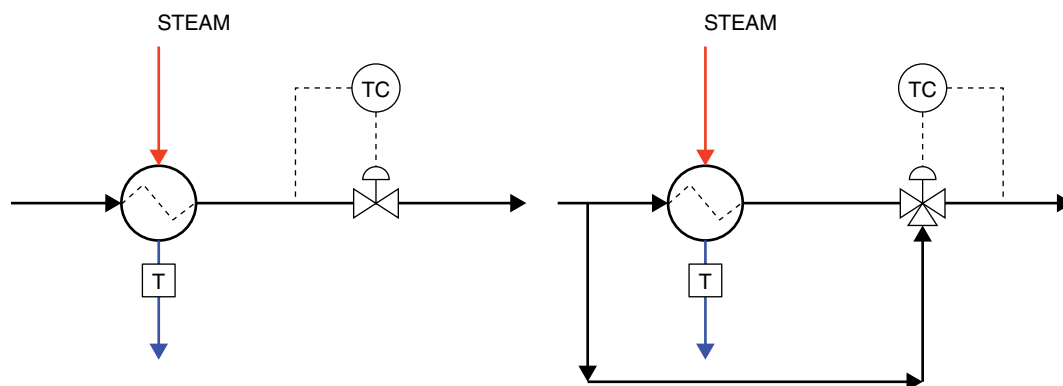


Figure 15.63 Methods of steam heater control.

Figure 15.62 shows two methods of steam heater control: direct control and bypass control. If you recall, we said that under normal circumstances, we would put the sensor on the target stream whose temperature we want to control, and the control valve should be on the other stream. However, in this example, because we want to avoid the complications associated with the phase-changing stream, we have installed both a sensor and a control valve on the non-phase changing stream, which is not a good design. Therefore the designs in Figure 15.63 are both pure control on a non-phase changing stream.

We can use direct control if a fluctuation in flow downstream is not important. However, if the equipment downstream needs a steady flow, then we need to opt for bypass control of the steam heater.

Bypass control can be used when flow changes caused by the control valve in the direct control method are not acceptable. However this control cannot satisfy us for all the cases. The bypass control means a portion of the stream is “cooked” and the other portion is “raw” and mixing these two streams together is not always acceptable.

A steam trap is installed on the condensate line to control the flow of condensate out of the steam heater. This serves to prevent steam from escaping from the system, which would be inefficient and uneconomical. We want to use all of the energy of the steam and convert it to condensate in the steam heater.

As we are not completely happy with the direct or bypass control of non-phase changing streams we are thinking about implementing the third option, partial flood control.

The purpose of a heat exchanger is to transfer heat from one stream to another. The amount of heat transferred is a function of the flow rate and temperature differences, but also of the surface area available for heat transfer. So, the higher the flow rate and the larger the

surface area, the higher the heat transfer between the streams.

We can blank off a section of the heat transfer area to decrease the amount of heat transferred in order to control the temperature. How can we blank off a portion of the heat transfer area? By covering it with a liquid; that liquid is the condensate form of the phase-changing medium. The reason that liquid coverage works to blank off the heat transfer area is that a liquid almost always has a much lower heat transfer coefficient than a phase-change heat transfer coefficient.

For instance, if we use steam as a heating medium in the heat exchanger, it will form condensate, which is drawn off. We can install a control valve on the condensate pipe and when this valve is closed, the condensate will flood back into the heat exchanger, covering part of the surface area and reducing the amount of heat transferred. This is the basis of control by partial flooding, which is shown in Figure 15.64.

If the plan is to use partial flooding, this should have already been discussed with the heat exchanger manufacturer. You cannot apply partial flooding in conventional heat exchangers. The other point is that, for this type of

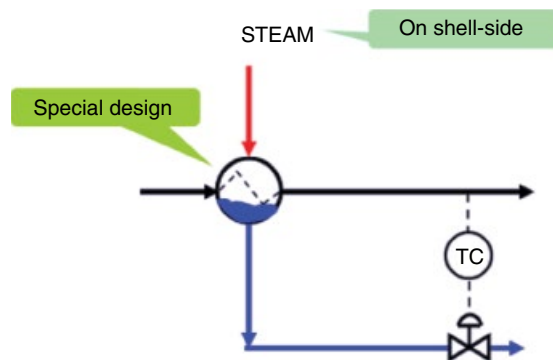


Figure 15.64 Heat exchanger control by partial flooding.

control, the phase-changing stream should be on the shell side of the heat exchanger.

Be careful: in this case, we don't use a steam trap, because we want our control valve, and not the steam trap, to control the condensate stream out of the heat exchanger.

The arrangement shown in Figure 15.65 will not work for control by partial flooding. If you pinch back the control valve, the steam trap will not allow the condensate to flood back into the steam heater.

You may say that partial flooding is not a good idea because it requires some changes inside the equipment and you prefer using "ordinary" methods for control of a heat exchanger with a phase-changing stream.

The alternative arrangement for "ordinary" control is shown in Figure 15.66.

Steam traps are not available in large sizes (possibly over 4 or 6 in.). If the condensate pipe is large, a condensate pot with level control can be installed instead. A steam trap is just a miniature version of a condensate drum with level control on it. However, in this case we need to be careful to locate the condensate pot far below the heat exchanger, otherwise there is a chance of

backflow into the heat exchanger. In such cases, basically we would have unintended, unwelcome partial flooding.

However, the above design has one main problem. The problem is – as you can see – that there are two control valves on the heating medium stream: one on the steam and the other one on the condensate. Two control valves generate a considerable amount of pressure drop. If the original pressure of the steam is not high enough, the system doesn't work because the steam cannot afford to travel from the steam header to the condensate header.

There are usually three types of steam available for use in plants: low-pressure steam, medium-pressure steam and high-pressure steam. Low-pressure steam is also named utility steam. The arrangement shown in Figure 15.66 will only work for high-pressure steam.

If the steam pressure is low (like in the case of utility steam), the arrangement shown in Figure 15.67 solves the problem.

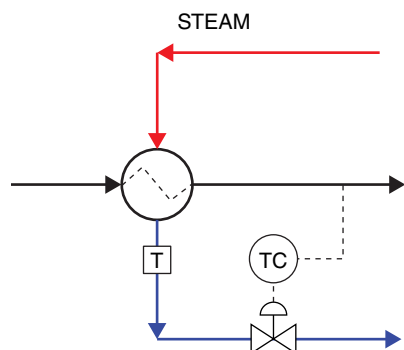


Figure 15.65 Poor control design.

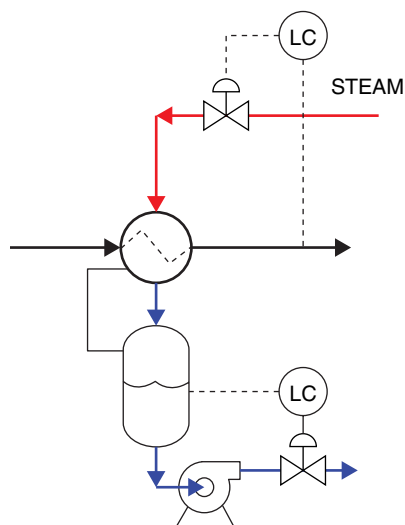


Figure 15.67 Steam heater control for utility steam.

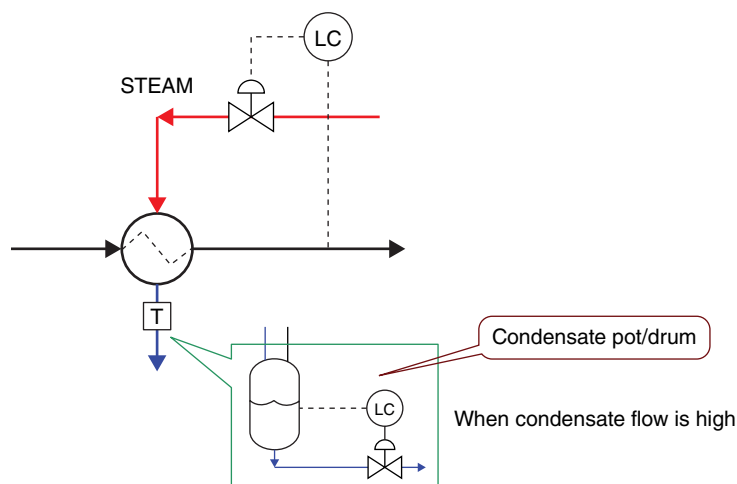


Figure 15.66 Steam heater control for high-pressure steam.

By installing a pump on the outlet of the condensate drum, we compensate for the pressure drop across the system and provide enough pressure for the operation of the control valve on the condensate line. This arrangement is applicable for utility steam and steam at any pressure.

An equalization pipe connects the heat exchanger to the drum to facilitate the removal of condensate from the heat exchanger. As you can see, if you try not to use a partial-flooding control system, there are some other problems you need to solve.

Now let's come back to the partial-flooding control system explained above. The above arrangements are shown only to introduce the concept of partial flooding. That arrangement, however, has some problems from a dynamic view point, which cannot be recognized from the P&ID. First, the temperature control loop is very sluggish. Second, if the temperature on the other stream drops, the control valve will open to reduce the surface area in the heat exchanger that is flooded, which will increase the heat transfer. However, if the control valve is opened too much, you face the danger of steam escaping from the system.

We can solve these issues by using the control arrangement shown in Figure 15.68.

By installing an elevation-adjusted condensate pot with level control, we can speed up the response time of the control system and ensure that no steam escapes. It is important to note that in order for control by partial flooding to work, the phase-changing stream should be on the shell side and in some cases the heat exchanger is installed vertically.

15.8.2 Air Cooler Control

The control of an air cooler is basically a combination of heat exchanger control and fan capacity control. These

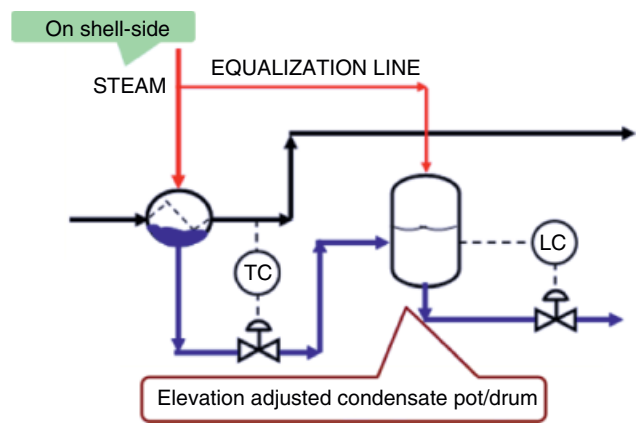


Figure 15.68 Steam heater with elevation adjusted condensate drum.

two separate loops are connected to each other in this way: a temperature signal from the outlet stream controls the fan capacity.

As was discussed previously, a fan's capacity can be controlled by suction throttling (here by changing the louvers), changing the rotation speed of the fan, and/or changing the blade pitch of the fan.

The schematic in Figure 15.69 shows a temperature loop that controls the louvers in the air cooler. The louvers act just like butterfly valves to control airflow.

In this type of control temperature of the target stream is adjusted by adjusting the louver on the air side.

Instead of control of louvers, the blade pitch (angle) of fan(s) can be changed based on an order from the temperature controller. The third option is changing the speed of the electric motor by VSD.

Figure 15.70 depicts a complicated air cooler control scheme. Here, all of the available control options are implemented through a split-control system. This is an uncommon type of split control. In the majority of cases, a split-range control system has only two ranges, but here we have three.

This three-ranged split range control could possibly be used where the temperature variation is high, but normally we would only control two of these variables using split range.

15.8.3 Heat Exchanger for Heat Recovery

If the purpose of a heat exchanger is heat recovery, it may not need any type of control. The reason is obvious,

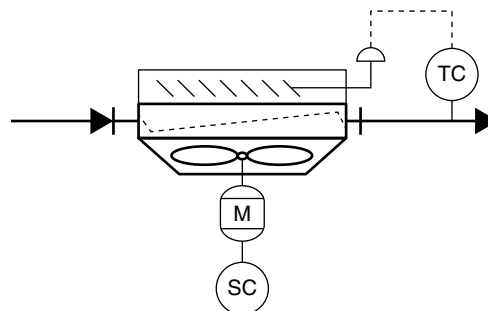


Figure 15.69 Air cooler control.

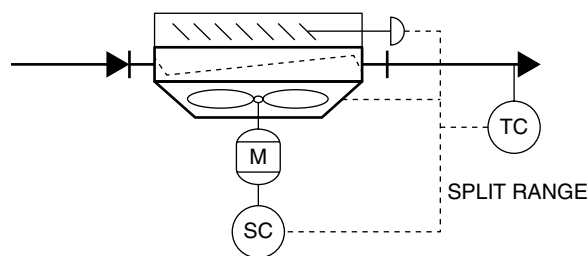


Figure 15.70 Air cooler with split-range control.

when the goal is transferring heat from one stream to another, we need to allow heat to be transferred as much as possible without any interference. Therefore no control is needed. These types of heat exchangers are seen a lot when trying to recover the heat as much as possible from the products of a unit to the feed of the same unit. The target stream as the giver of energy is identified and its heat is transferred to a cold stream. It may be asked: is the temperature the cold stream left uncontrolled? The answer is if the temperature the cold stream is important, another heat exchanger (most likely a utility heat exchanger) needs to be placed on that with control system on it.

15.8.4 Back Pressure Control of Heat Exchangers

Some heat exchangers may need an additional control to make sure their temperature control works efficiently. Not all heat exchangers need this additional control but there are cases that they need this, but it was overlooked when the P&IDs were developed and it causes problem in operation.

As we know, heat transfer rate in heat exchangers are a function of flow rates (in addition to other parameters). If a heat exchanger is placed on a pipe with “non-suitable” hydraulic features, the temperature control loop cannot control the heat exchanger. This means we need to make sure we have enough flow rate in the heat exchanger to make the temperature control loop fully functional.

In Figure 15.71 there is a heat exchanger on a stream that comes from very high pressure (on the upstream side) and goes to an atmospheric tank (on downstream); stream A. The temperature control loop is that control valve is installed on the other stream; stream B.

You may guess that the flow of stream A could be huge. When a huge flow rate goes through the narrow tubes of a heat exchanger the velocity of flow could be so high that heat transfer cannot be happen efficiently. Therefore the flow rate of stream A should be controlled to make sure a good heat exchange happens in the heat exchanger.

This can be done by controlling the pressure of stream A on the downstream of the heat exchanger. Controlling this pressure could be done by a pressure control loop or a back pressure regulator.

The set point of control loop could be adjusted on a pressure lower than the heat exchanger upstream pressure, but not very low.

15.8.5 Fired Heater Control

A fired heater is also a type of heat transfer operation. The control of a fired heater comprises the four following elements:

- 1) Coil outlet temperature (COT)
- 2) Firing control
- 3) Skin temperature control
- 4) Pass balance control.

COT control is the primary goal of a fired heater. Similar to heat exchangers, here again we have to control the temperature of the target flow. COT control works closely with the second element, firing control, to make sure we have enough heat released into the fired heater chamber.

Skin temperature control is needed wherever there is a fire (burner) in process equipment. Because there is always some uncertainty in the length and shape of flames there could be always the chance of flame impingement with the heater internal coils. The flame impingement may cause coil burning and rupture. To prevent this, the temperature of the external surface of the coils, the “skin,” should be sensed and controlled. Sometimes such control is only of SIS type, and some others BPCs type, or both.

The fourth control is nothing more than controlling flow in parallel pipes. As mentioned in Chapter 11, each fired heater may warm up a stream, not through a single coil, but rather through several coils. The flow splitting must be done in such a way that the flow is evenly divided among the different passes. This is the concept of pass balancing. As you may guess, this practice can be done using the concepts of flow splitting and merging, which were discussed previously.

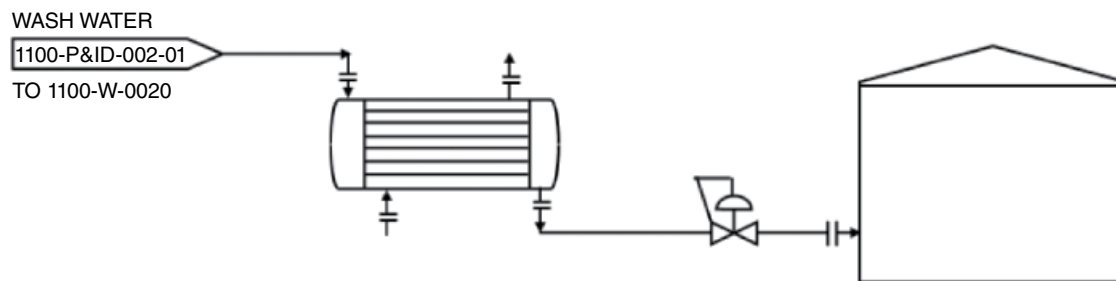


Figure 15.71 Backpressure regulator on the outlet of heat exchanger.

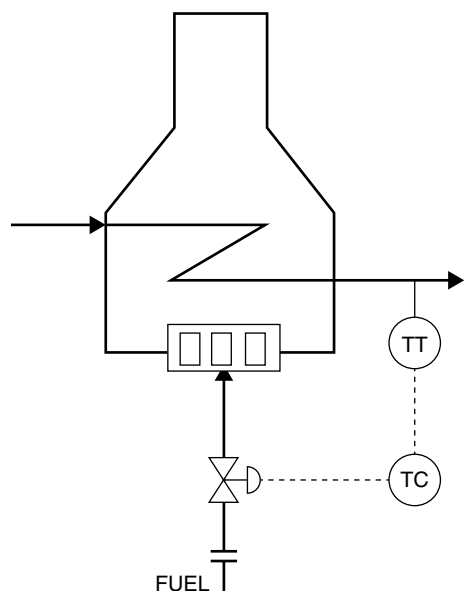


Figure 15.72 Simple fired heater control.

Figure 15.72 shows a different example of control on a single-pass fired heater. The firing control schemes for these examples are very simple; we will discuss more complicated firing control schemes later in this section.

This schematic shows the simplest method of control by using a temperature loop to adjust a control valve on the fuel supply line to the heater.

As we know, a temperature control loop is sluggish. To speed up the response time, we can cascade the temperature over the flow, where the temperature loop provides an RSP for the flow controller on the fuel line, as shown in Figure 15.73.

If the fuel is a gas (for example, fuel gas), the flow loop can be replaced with a pressure loop.

The example in Figure 15.74 shows temperature control of a fired heater when the temperature of the process fluid swings a lot and the burner may start to reach its overly high end. In that case, the heater coil needs to be protected from burning out.

The normal operating, or acting, loop measures the temperature on the discharge line from the heater to control a valve on the fuel inlet. As we know, a temperature loop is very sluggish, so we install a faster loop to override the normal operating loop if the temperature drops too low. The override loop is connected to a temperature sensor in the fired heater.

The next example shows a more complete control scheme for a fired heater. In this example (Figure 15.75), we have included the control of airflow as well.

First, we have a temperature loop controlling a valve on the fuel line. Then we have a flow loop, with sensors on both the fuel and air lines. We want the airflow to follow the fuel flow according to a specific ratio. The

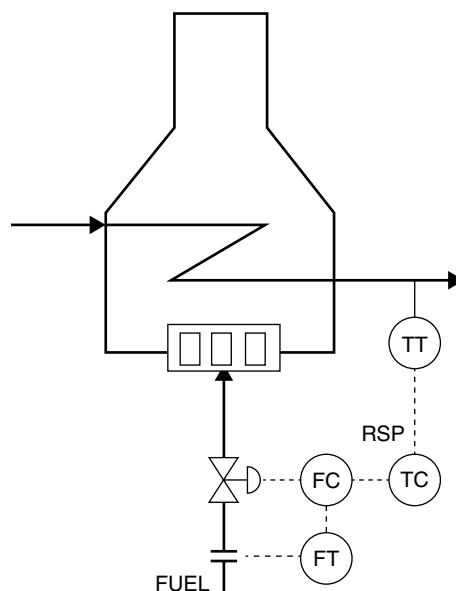


Figure 15.73 Fired heater with cascade control.

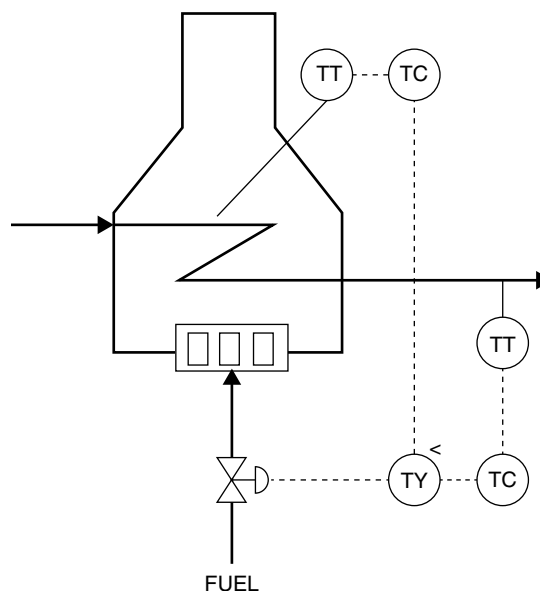


Figure 15.74 Fired heater with override control.

flow sensor on the fuel line provides an RSP for the FFC. The FFC receives a signal from the flow sensor on the air line as well, and produces a new signal according to the ratio setting to control the valve on the air line.

In the next method, shown in Figure 15.76, we have a similar action as in Figure 15.75; here, however, the flow sensor is on the “flue gas” stream and not the “air stream.”

We cannot put a flow sensor on a stack because it is too big. We solve this by using a DP (differential pressure) measurement from two points on the stack. The DP is directly proportional to the flow.

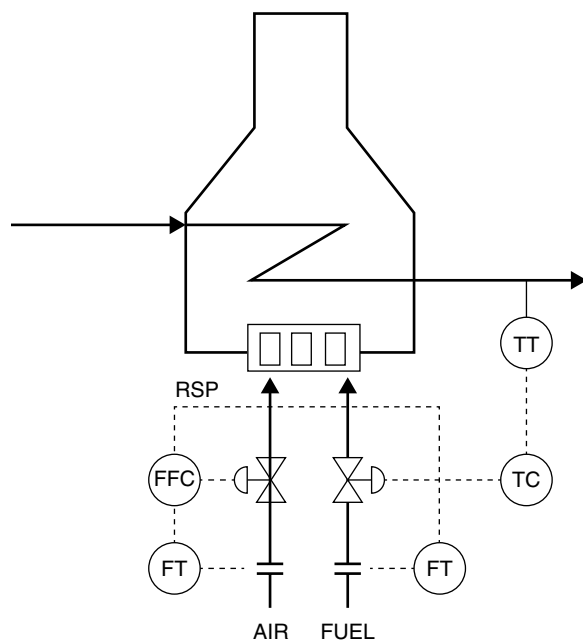


Figure 15.75 Fired heater with ratio control of airflow.

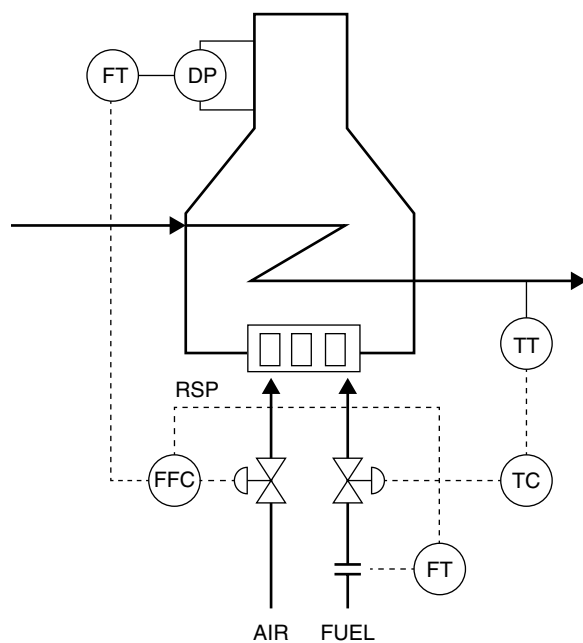


Figure 15.76 Fired heater with ratio control.

The airflow is not necessarily controlled as a ratio of fuel flow. It can be controlled independently based on residual oxygen in the flue gas. Firing in a fired heater is always done by “excess air.” This means that more air must be provided than what is stoichiometrically required by the fuel. This is to make sure the flame always receives enough oxygen for burning. Based on this concept, we can control the combustion air (air required for

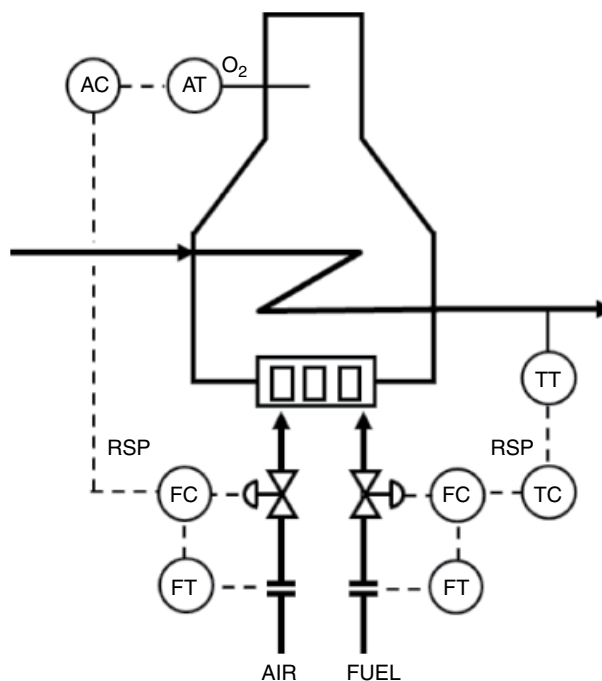


Figure 15.77 Fired heater with residual oxygen control on the flue gas stream.

combustion) based on unburned oxygen in the flue gas stream.

This type of control is shown in Figure 15.77.

Now we are going to move on to more complicated firing control schemes. These types of control are generally for the cases where our fuel is fuel gas. Generally speaking, liquid fuels don’t have the problems that require the use of more complicated control systems.

In all of the above examples, we actually assumed that “by controlling the fuel gas flow rate we are controlling the heat released in the fired heater.” However, this assumption is not always right. The amount of heat that can be released from the fuel gas upon combustion is called the “fuel heating value.” If the heating value of the fuel gas is constant, the above assumption is true. That is the case for a fuel gas that comes from a pipeline and is bought from a fuel gas company, which has a fairly fixed composition.

Sometimes, however, the fuel gas has a changing composition, so its heating value is also changing. This could be the case when you are using one of the plant’s gaseous process streams as a fuel gas. When using an in-plant stream as fuel gas, there is a high chance of having a variable heating value. In such cases, we cannot say that by controlling the fuel gas flow rate the released heat can be controlled. In such cases, we need to find more complicated ways to “control the incoming heating value” to the fired heater.

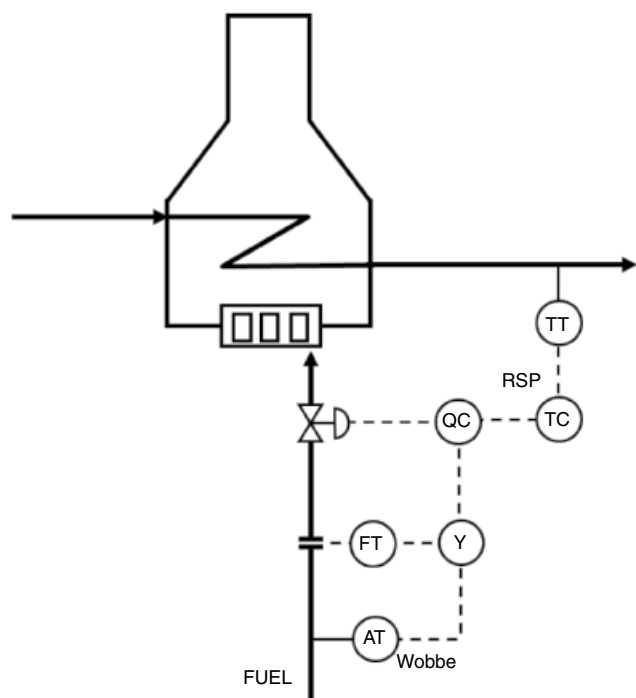


Figure 15.78 Firing control of a fired heater when the heating value of fuel gas fluctuates severely.

The silver bullet for the firing control of a fired heater is to use a type of heating-value meter to sense the heating value of the incoming fuel gas in a timely fashion. One type of these meters is the “Wobbe index analyzer.”

A Wobbe index analyzer, together with a flow meter, provides the heating value rate (e.g. in BTU/hour). A control loop based on a “Wobbe index analyzer + flow meter” can do the job, or, if we are looking for a speedy response, a cascade control of coil temperature over “Wobbe index analyzer + flow meter” does the job too.

The schematic in Figure 15.78 shows a cascade firing control system based on a Wobbe analyzer.

For cases where the heating value of the fuel gas doesn’t fluctuate a lot, it can be assumed that the heating value of the fuel gas is a function of the fuel’s molecular weight. There is no online analyzer for continuously measuring molecular weight; however, as the molecular weight is a function of specific gravity and volumetric flow rate, the molecular weight can be monitored by an online specific gravity analyzer (“SG analyzer”), and the fuel gas flow rate.

This type of control is shown in Figure 15.79.

If the fluctuation of the fuel gas composition, and therefore its heating value, is very limited, it can be assumed that the heating value of the fuel gas is a function of the fuel gas mass flow rate. In this case, a control loop based on a mass flowmeter can control the firing of

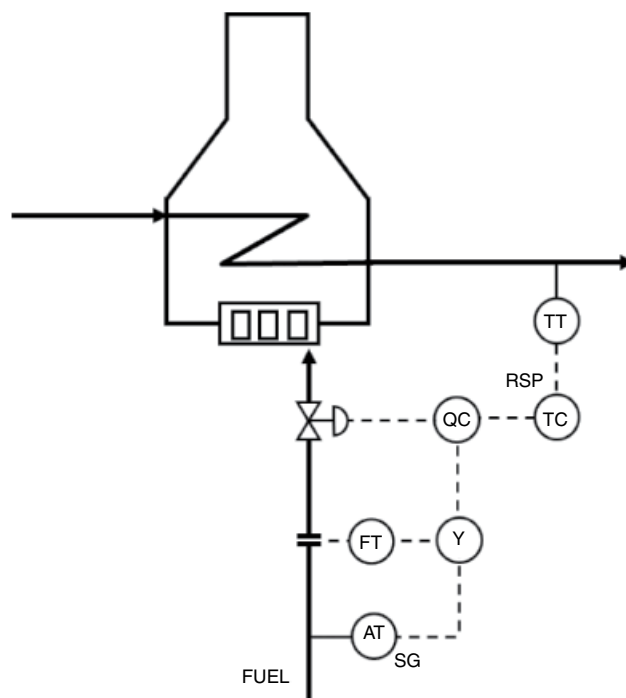


Figure 15.79 Firing control of a fired heater when the heating value of the fuel gas has limited fluctuation.

the fired heater. Therefore, the control here is very similar to the simple control that we had for the constant-composition fuel gas, except that the (volumetric) flow meter is replaced with a mass flowmeter.

All of the control systems described above applied to what we call “once-through fired heaters.” Once-through fired heaters are a type of fired heater wherein the process fluid only passes through the inside of tubes.

There is one other type of fired heater where the process fluid goes through it within a bundle of drums and tubes. This type of heater is almost exclusively used as a boiler. This boiler is called a “drum boiler.” Drum boilers need more control loops than once-through fired heaters. On additional control loop on drum boilers is the “steam drum level controller.”

15.9 Container Control System

Container control is discussed in full in Sections 15.4.5 and 15.4.6 of this chapter. The text below is a brief summary of that discussion.

Whenever we have a liquid or a gas in a system, we like to know how much we have. For all gas containers, we can control the inventory with a pressure loop. For non-flooded liquid containers, we control inventory by using a level loop.

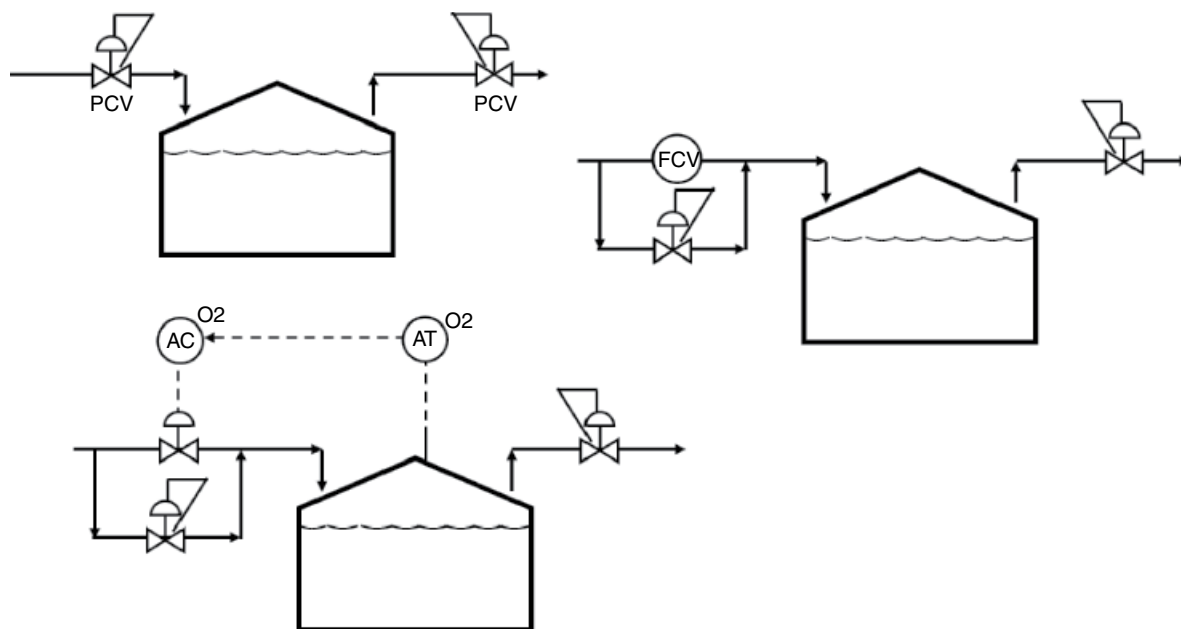


Figure 15.80 Blanket gas control systems.

If we have liquid in a flooded container, we don't really need inventory control. A flooded container with liquid can be pictured as a piece of pipe! That is why, in some instances, such as in industrial water treatment plants, we don't have many control loops. This is because the majority of the vessels in the plant are flooded, and one control loop may be able to control a long string of units.

If a container is non-flooded, the liquid phase (on the bottom) and the gas phase (on top) can be controlled independently. A "blanket gas control system" is an example of such cases.

15.10 Blanket Gas Control Systems

Blanketing is used in non-flooded containers (generally tanks) for different reasons. However, the goal is always to provide a positive pressure of a gas of choice in the empty space of the tank. This positive-pressure gas works similar to a cushion or pad on the surface of the liquid inside of the tank. There are different control systems available, and they are shown in Figure 15.80.

In the top-left arrangement, we have pressure regulators (PCVs) on both the inlet and outlet blanket gas

streams. You can picture a pressure regulator as being a pressure loop inside of a casing. A regulator is a mechanical type of control loop that takes the place of a pressure loop.

The pressure regulator on the inlet blanket gas stream is a conventional pressure regulator, while what we have on the blanket gas outlet is a "back pressure regulator." In a conventional pressure regulator, the pressure sensor is located downstream of the control valve, but for a back pressure regulator, the pressure sensor is upstream of the control valve.

Another arrangement (top right) uses a flow regulator (FCV) on the inlet line. An FCV is also a type of regulator, but the sensing parameter is flow rather than pressure. So, it can be said that "a flow regulator is a flow control loop inside a casing."

The third blanket gas control scheme is shown on the bottom of Figure 15.80. In this arrangement, we focus on the main purpose of using blanket gas, which could be the prevention of the exposure of the process liquid to oxygen. In this case, we install a process analyzer to check how much oxygen we have in the tank. Instead of a pressure regulator on the inlet, we have a composition loop to control the valve.

Reference

- 1 Woodside O. Protect centrifugal pumps from low flow, chemical engineering progress, June 1995, p.53

16

Plant Interlocks and Alarms

16.1 Introduction

In this chapter we cover SISs, alarm systems, discrete control, and electric motor control.

SISs and alarm systems are two layers of steering processes that come after BPCS.

Discrete control is a type of BPCS control, but it is covered here because its actions are similar to SIS actions.

Electric motor control, including through BPCS and SIS, will be discussed as an important example later.

16.2 Safety Strategies

It is mentioned in Chapter 13 that “taming” process parameters is mainly done through the concepts of BPCSs and SISs. When the control of a parameter passes from the hands of the BPCS to the SIS, an alarm is raised to warn the operators.

As mentioned in Chapter 12, there are four methodologies to cope with safety issues: inherent design, passive action, active action, and procedural action.

A SIS is set of active actions implemented in process plants. A SIS is a highly regulated component of process control because of its effective action of mitigating safety issues. In addition to codes generated by regulatory bodies there are well-known standards regarding SISs that can be agreed upon to be followed in a process plant. Instrumentation and control practitioners are the professionals looking after such issues and in this chapter there is no intent to provide a complete approach to SISs.

16.3 Concept of a SIS

The concept of a SIS is shown in Figure 16.1.

The impact could be because of operator error, the process went out of control, equipment failure, power loss or other things.

After the impact, one or more process parameters go beyond the safe operating range.

In the last step one or more functions of SIS are triggered to bring the process back into its dedicated “playground.”

A SIS is basically a set of SIFs (safety instrumented functions) while a PBCS is a set of control loops.

Essentially, a SIS does whatever an operator is supposed to do when an alarm sounds.

16.4 SIS Actions and SIS Types

SIS actions are mainly two types: action on switching valves (on or off) and/or actions on rotary machines (start-up or shutdown).

A SIS triggers one or more SIFs to mitigate the impact. If there is more than one SIF to be triggered they could be “connected” to each other by logical operators like “AND” or “OR” or other more complicated logic.

A SIS action band is between a BPCS band and a mechanical relief band. Thus the first objective of a SIS can be defined as “bringing a stubborn parameter back to the BPCS band.”

The other point here is that triggering a mechanical relief system is not a good thing. It is because consequences of each time functioning of a mechanical relief displacing the process fluid from inside of process units to other part of “systems” which is not a good action.

If the mechanical relief system is the overflow pipe of a tank, when it works, the liquid comes to the outside area, which is not a pleasant event and could be dangerous, depending on the type and temperature of the liquid.

When a mechanical relief system is a pressure safety valve (PSV), it may release to the atmosphere or an emergency release system like a flare. It could be said these systems are designed for such emergency release

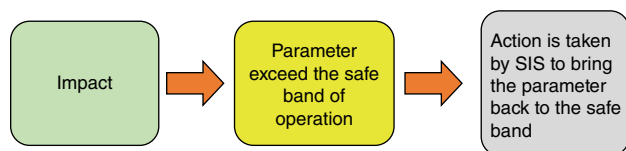


Figure 16.1 Concept of SIS.

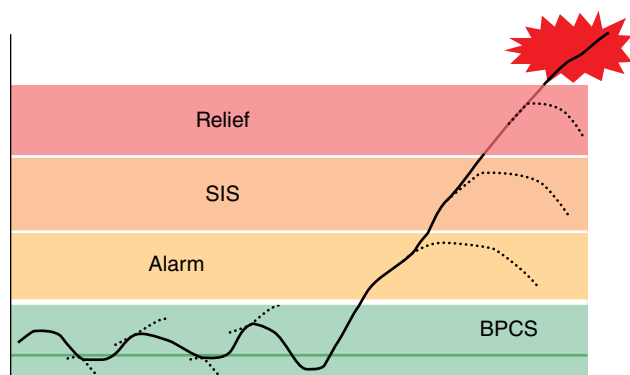


Figure 16.2 SIS band.

and everything is still under control even after popping the PSVs. However, it is not right, especially if there is a big emergency and several PSVs pop at the same time.

Therefore the second duty that can be defined for a SIS is routing the “panicked process fluids” to a route other than the mechanical relief (or PSV).

The third SIS duty is defined as limiting the area of the emergency and preventing it from expanding to other areas of the plant (Figure 16.2).

The actions of a SIS are any of, or combination of, the below actions:

- The actions to prevent each parameter from triggering the action of a mechanical relief system (e.g. pressure safety valve).
- The actions to minimize the amount of released process fluid through a mechanical relief system (e.g. pressure safety valve).
- To get rid of a mechanical relief system (e.g. pressure safety valve) or to decrease the size of it (and consequently the size of emergency collection network).

In the next paragraph it can be seen that a SIS for each of above goals has a specific name.

SISs are of different types. A SIS is the collective name for different safety-related systems in plants. A SIS function can be classified into one of the following classes, depending on the goal of the SIS function:

- ESD. “Emergency shutdown.”
- Emergency isolation. In case of fire in the plant, you need to isolate one or more areas that are at particular

risk from fire, such as an area containing flammable liquids.

- Emergency depressurizing and blowdown. For example, in the case of a fire, we need to reduce the risk imposed by high-pressure gases, by depressurizing and blowing down to get rid of the gas.
- HIPPS. The “high-integrity pressure protection system” is to protect equipment against overpressure.
- BMS. A “burner management system” is a safety management system for the start-up, operation and shut-down of burner units such as boilers and fired heaters, which may use any type of burners in their system.

An ESD system is arguably the major constituent of a SIS. Some people, when they think about SISs, have only an ESD in mind, which is not correct.

The backbone of an ESD is: “minimizing the energy in the system.” This can be done mainly by:

- Minimizing the temperature. For instance, you can open a switching valve on a cooling water circuit to decrease the temperature, or close a switching valve on fuel gas to burners, etc.
- Minimizing the pressure. This can be achieved by opening a switching valve on a vent pipe.
- Minimizing the number of moving mechanisms. We usually try to shut down rotating mechanisms such as motors that are connected to pumps and compressors.

An ESD is needed in process plants dealing with high pressure fluids and/or flammable materials.

An “emergency isolation system” may come after the ESD. If the closing off action of the ESD is not enough, and even small amount of leakage after closing the valves is still problematic, the isolation should be added to the system. Emergency isolation system is generally double block and vent/bleed switching valves.

In some more critical services where shutting down a flow is very critical, the double block and vent/bleed arrangement is used. This is an arrangement similar to the double block and bleed system for the purpose of isolation discussed in Chapter 8; however, everything here is automatic.

Figure 16.3 shows a schematic of such an arrangement.

“Emergency depressurizing and blowdown” are a type of SIS in which by opening some routes the contained liquid or gas is removed from the process equipment. Depressurization is referred to gas removal while blowdown means liquid removal.

Emergency depressurizing is common when dealing with some full gas containers. In full gas containers the pressure safety valves may fail to function at an appropriate time.

Figure 16.3 Automatic double block and vent/bleed switching valves.

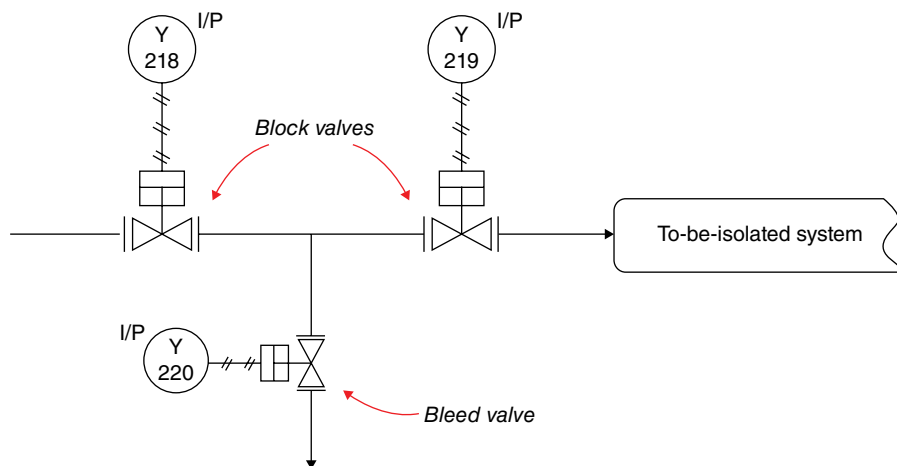
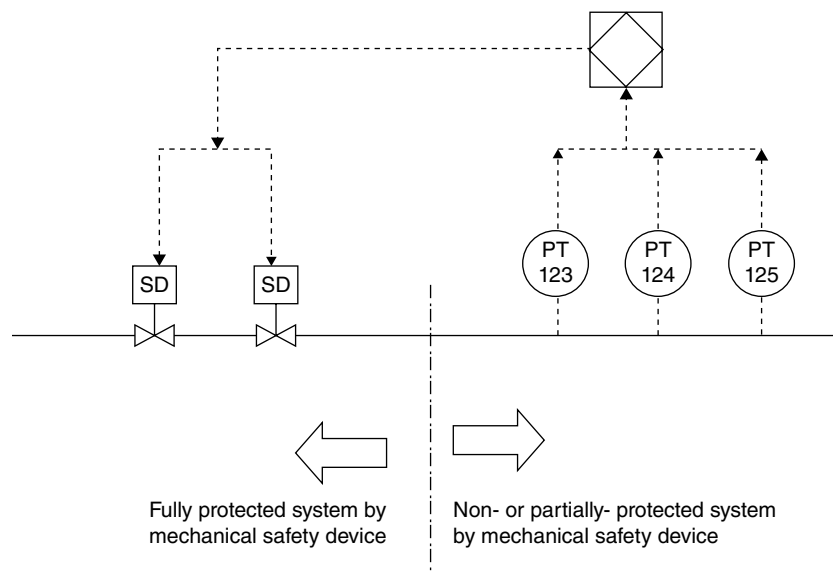


Figure 16.4 A typical HIPPS.



A HIPPS is a type of SIS that cuts off a high pressure, rebel stream to stop it from getting into a non- or partially protected system (Figure 16.4).

Implementing a HIPPS is highly regulated because it may release the need of pressure safety device downstream of HIPPS. HIPPSs are used to get rid of a pressure safety device or decrease its size.

A HIPPS is used where one or more pressure safety devices has such large release rate that severely impacts the collection network and/or emergent release disposal system. The other cases in which a HIPPS may be needed is for the cases where putting a pressure safety valve is not technically doable, like under-the-sea facilities.

The result of implementing a HIPPS on the downstream system could be reduced size of PSV,

removing PSV, or reducing the design pressure of the downstream system.

A BMS is a type of SIS that most likely exists when dealing with burners. Burners could be in a fired heater, boiler, steam generator or even a tunnel dryer. A BMS is a regulated practice in many countries. Wherever there is a burner in the system, a BMS needs to be implemented. A BMS includes actions similar to ones in an ESD and emergency isolation plus additional actions to push the potential accumulated flammable mixture out of the system. The BMS actions could be initiating snuffing steam and/or compressed air into the firing box.

A BMS could be a combination of other SISs like an ESD, emergency isolation, etc.

BMS is a collective name for different practices, but just few of them are visible on P&IDs.

16.5 SIS Extent

Ultimately, the objective of a SIS is to hold the plant within a safe “window” by applying safety strategies, both plant-wide and for individual equipment. This will ensure that the plant operates safely.

Each SIS function has a specific range of effect. There are at least four levels for a SIS function’s effectiveness range:

- Tier 1: equipment shutdown
- Tier 2: system shutdown
- Tier 3: unit shutdown
- Tier 4: plant shutdown.

Tier 1 affects only one piece of equipment. Tier 2 affects a system: a few pieces of equipment, generally related to each other. Tier 3 affects a unit: several pieces of equipment that work together to attain a specific goal. Tier 4 shuts down the whole plant.

The whole purpose of tier 1 shutdown is to handle the situation in a way that prevents the upset from “rippling” to the next level that may trigger tier 2 shutdown, and then tier 3 and tier 4.

It is important to know that definition of these tiers is only in the process engineering world, and is not cut and dried, and does not affect the software or hardware of their respective SISs.

16.6 Deciding on the Required SIS

Even though this is the last section on SIS functions, it could be considered as the most important. In this section, we answer the following question: “how does a process engineer decide to install a specific SIS function for a specific piece of equipment/point/area in a process plant?”

There are two methods that can be used to implement a SIS in a plant during the design phase:

- 1) Qualitative method. This method draws largely on previous experience and the details are normally verified later in a HAZOP meeting. HAZOP is discussed at the end of this section.
- 2) Quantitative method. This is the more reliable method of designing a safety system. This is done by performing a SIL (safety integrity level) evaluation. This evaluation is conducted in a SIL review meeting, where a particular action or piece of equipment is assigned a SIL rating from one to four. A rating of four is critical and means that a safety system has to be implemented.

In reality, it would be very time consuming to do a complete quantitative analysis of the plant and all the equipment through a SIL review. Most people tend to

use a combination of the two methods, starting with a HAZOP meeting. What can’t be decided in the HAZOP meeting – hopefully just a few items – goes to the SIL review meeting for quantitative analysis. Based on this and a LOPA (layer of protection analysis), we can decide whether we need another layer of protection around the system or not.

To give you an idea of SIL ratings, nuclear systems are usually assigned a SIL of four. In the chemical industry, three. Some companies accept merging the SIS loop of any specific event with a BPCS loop if the SIL rating of an event is one or two.

16.7 The Anatomy of a SIS

The general schematic of the SIS concept is shown in Figure 16.5.

The first element of a SIS is a sensor, which could be for level, flow, temperature, pressure or composition. That said, we try to avoid including a composition sensor in the system because it is not very reliable. The sensor sends a signal to a logic solver, which is usually a PLC. From there, a signal goes out to the actuator, which could be a switching valve, or the on/off on a rotary machine.

Sensors are set to trigger a SIS action automatically, but we can also make provisions for pushbutton intervention. With this feature, the operator can shut down specific equipment, or the whole plant, through a “manual SIS.”

The sensor and actuator in a SIS need to be located close to each other – much closer than you would find in a BPCS. This is because when the system senses a predetermined danger point, it needs to be shut down as soon as possible.

The sensitive points could be high-high, low-low, high or low. We usually place a SIS action on the high-high and/or low-low positions; high and low positions are generally “reserved” for alarms. As we know, there should be enough time (between 5 and 30 minutes) between the alarm and the SIS action, to give the operator time to take corrective action before the SIS activates.

16.7.1 SIS Element Symbols

As we learned, there are three main elements in each SIS function: the primary element, the final element, and the logic solver, or the “brain” of a SIS function. The P&ID symbology for each of these is discussed below.

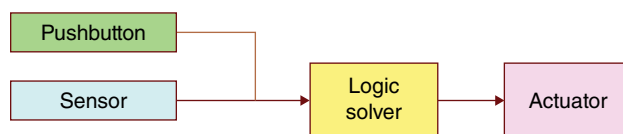


Figure 16.5 SIS anatomy.

16.7.1.1 SIS Primary Elements: Sensors

The primary element of a SIS function is a sensor, which is sensitive toward specific points, high-high, low-low, high, or low. Therefore, their acronyms start with the process parameters they are sensitive to: pressure, flow, level, or temperature, followed by the sensitive point: HH, LL, H, or L.

You can see a list of different SIS primary elements in Table 16.1.

For example, PEHH means a pressure element that is triggered on a high-high setting. FEH means a flow element that will activate on a high setting. These tags don't have any divider because they are usually in the field.

In addition to process parameters, non-process parameters play a vital role in a SIS. The non-process parameters are the parameters other than flow, level, pressure, temperature, and composition. These parameters are generally related to equipment rather than process materials. Examples are vibration parameters from compressors and centrifuges, and torque from rotating blade of sedimentation basins. Non-process parameters are less important in a BPCS rather than a SIS. The reason is that non-process parameters are not considered as parameters very close to the process and their change is not closely related to process.

In the older days, we didn't use sensors for SIS functions; instead, we used "switches." Switches were a very reliable type of sensor that were sensitive only toward a point, rather than to a range. In those days, we used

acronyms like "PSHH" to show a "pressure switch which is sensitive to the high-high point." These days, sensors are as reliable as switches in that they can be used both in BPCS loops and in SIS functions. Table 16.1 shows the features of switches and sensors.

16.7.2 SIS Final Elements

SIS final elements are the elements that "administrate" the action determined by the SIS logic for the process. SIS final elements are generally one of the below two items:

- Electrical motor switch
- Switching valve.

You may notice that these two elements are devices for initiating and stopping flows; an electric motor connected to a fluid mover initiates a flow and a switching valve can stop a flow by closing a pipe.

16.7.2.1 Switching Valves

A switching valve is a remotely operated on/off valve. In Figure 16.6 "SD" stands for shutdown.

This valve is not throttling and has just two positions: open or closed.

Figure 16.7 shows a switching valve, a remotely operated on/off valve, which gets its orders from a process parameter.

The schematic in Figure 16.7 shows a piston-type switching valve that is normally activated by a pneumatic signal. However, the signal coming from the controller is electrical, so we need a converter or transducer to convert it to a pneumatic signal. LY shows the transducer.

L means the process parameter is level but it could be any other parameters of P as pressure, T as temperature, F as flow rate, or dozens of analyte parameters as A.

Y is a non-defined letter, therefore it is defined outside of the circle as I/P, means "electrical current to pneumatic signal converter."

The transducer is placed in a circle without a divider line as it is generally in the field.

A sequence number specifies the SIF number, which should be unique for all elements of the SIF.

Switching valves in SIS functions can be tagged in different ways:

- XV – the X indicates that the valve can be activated by a signal from one event.
- UV – the U indicates that the valve can be activated by a signal from more than one event.

You may have seen switching valves that are tagged as KV, but it is important to understand that these do not have SIS functions. In a KV tag, the K indicates that the valve can be activated based on a specific time lapse, but

Table 16.1 SIS primary elements.

	HH	H	L	LL
P	PEHH 342	PEH 342	PEL 342	PELL 342
F	FEHH 342	FEH 342	FEL 342	FELL 342
L	LEHH 342	LEH 342	LEL 342	LELL 342
T	TEHH 342	TEH 342	TEL 342	TELL 342

Table 16.2 Features of switches and sensors.

	Sensor	Switch
Application	Regulatory control loops	Safety instrumented functions
Target sensing	A range	A spot
Reliability	Less reliable in older days	More reliable

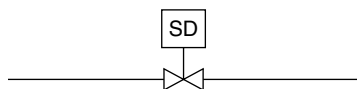


Figure 16.6 Switching valve as SIS final element.

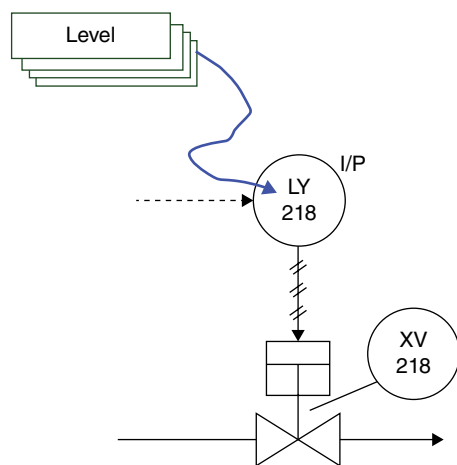


Figure 16.7 Switching valve.

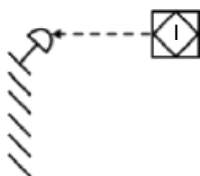


Figure 16.8 Louver as SIS final element.

generally there is no safety function that works based on time. You may find switching valves with a KV tag in discrete control loops. Discrete control loops are discussed in Section 16.9.

The structure of switching valves is discussed in Chapter 7. To refresh your memory a summary is given here.

A switching valve consists of an on/off-type valve (like a gate valve or ball valve), plus an on/off actuator. Different types of on/off actuators are available, including pneumatic actuators (diaphragm or piston type), hydraulic actuators, motor actuators and solenoid actuators.

One other device that can be considered as a SIS final element is a “remotely operated louver.” Figure 16.8 shows the symbols of a remotely operated louver in this application.

16.7.2.2 Switching Valve Actuator Arrangements

In some P&IDs, whenever there is a switching valve, the actuator arrangement of the switching valves is shown, in addition to the symbol for the switching valve. However, this makes P&IDs more crowded.

A switching valve can be either spring loaded or double acting, as shown in Figure 16.10. During normal operation, a spring-loaded valve using instrument air defaults to the open position, and it has a spring to keep it open. We call this “fail open.”

If the solenoid valve in the actuator is a three-way valve, the actuator is spring loaded. If it is a four-way valve, the actuator is double acting. These three- and four-way valves are not like process multi-port valves. They are used merely to direct the flow of instrument air; in the case of the four-way valve, the instrument air is directed either above or below the piston of the process control valve, or to vent.

Figure 16.9 shows the detail of an actuator arrangement for two types of actuators, spring loaded and double acting.

16.7.2.3 Valve Position Validation

Generally when we “communicate” with switching valves, we only send orders to them and ask them to close or open, but we really don’t know whether the switching valve actually carried out our orders or not.

Sometimes a valve becomes jammed in one position, for example in the closed position, and when we ask it to open, it won’t. So, in some more critical SIS functions (like switching valves in a BMS), we need a system to check what position the valve is in. This is done through the use of limit switches, marked ZSO and ZSC in the right hand loop in Figure 16.10, which are attached to the valve stem. O indicates open and C closed. The Z is used to indicate location.

The signals from the limit switches go to the indicator, XI, which should be situated in the control room. This allows the operator to check whether the valve is open or closed.

16.7.2.4 Merging a Switching Valve and a Control Valve

Can we merge a switching valve with a control valve? This is a very tempting idea to save some money. You may say, “there are two loops side by side, and each of them has a dedicated sensor and final valve. One of them is a control loop and the other one is safety logic. Can we merge them together?”

Many people believe that it is not a good idea to merge them, as it might compromise the integrity of the safety action. For this reason, it is not popular.

In general, it is good practice to keep SIS actions separate from a control loop (BPCS loop) to ensure plant safety. However, it is more economical to merge these valves and for this reason some people will do so if it doesn’t have a critical impact on safety.

However, if, for whatever reason, we decide to merge a control valve with a switching valve, how can we do it?

Figure 16.9 Actuator arrangements.

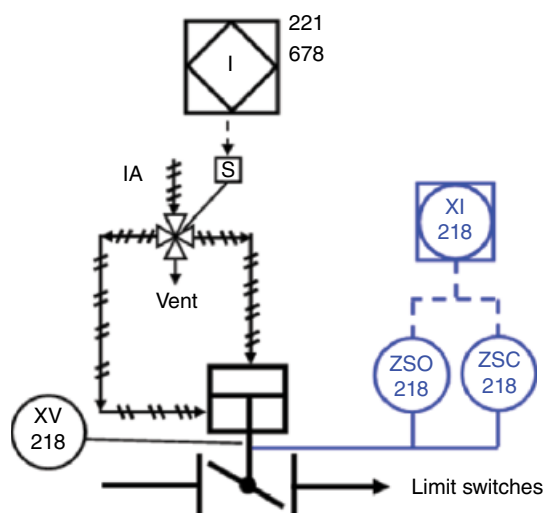
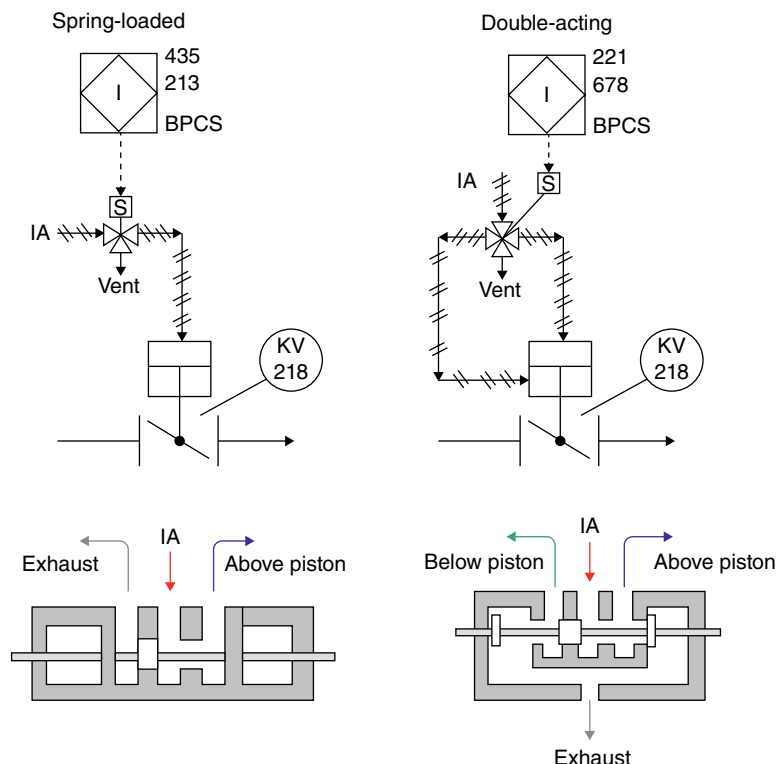


Figure 16.10 Valve position validation.

There is no problem with the sensor, because the same sensor is used in both cases. The schematic in Figure 16.11 shows a combined control valve + switching valve.

On the far left-hand side of the schematic, you can see a control valve. In the control valve, a transducer (or operator) converts the electrical signal (coming from the controller) to a corresponding pneumatic signal.

The transducer basically converts the instrument air (IA) pressure to a signal with pressure range of 3–15 psig.

In the middle schematic showing a switching valve, the instrument air signal, which is a single pressure signal, goes into the switching valve. A specific arrangement of solenoid valve(s) will continue or discontinue this signal per the orders received from the SIS logic.

To “build” a control/switching valve, it is only needed to add the solenoid arrangement to the IA input of the control valve (right-hand scheme).

16.7.2.5 On/off Action of Electric Motors

Electrical motors could be an electrical motor of a pump or a compressor.

The schematic in Figure 16.12 shows P&ID of a simple SIS on an electric motor driving a pump.

The diamond-shaped tag indicates a SIS action.

SIS actions on an electric motor will be explained in more detail in Section 16.12.

16.7.3 SIS Logic

The SIS function is handled by a “safety PLC.” As mentioned in Chapter 13, a PLC is represented by a diamond-shaped symbol. Therefore, the specifying characteristic of SIS logic is a diamond on P&IDs.

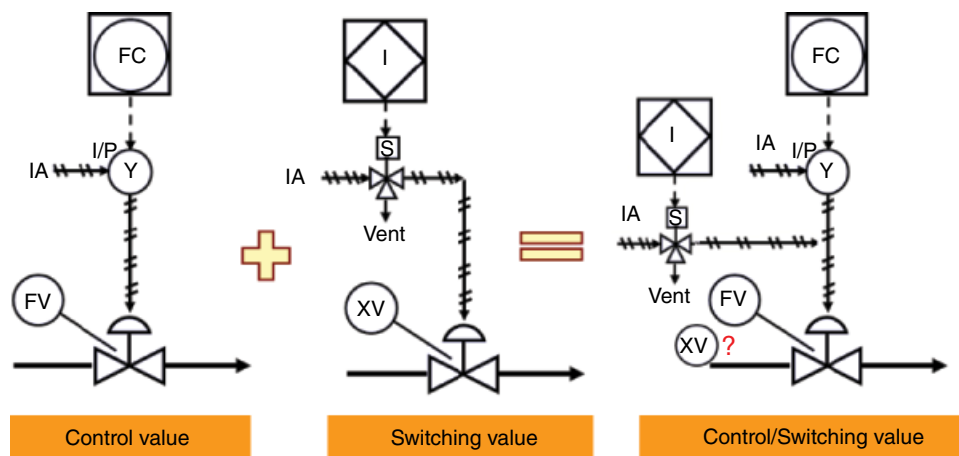


Figure 16.11 Merging switching and control valves.

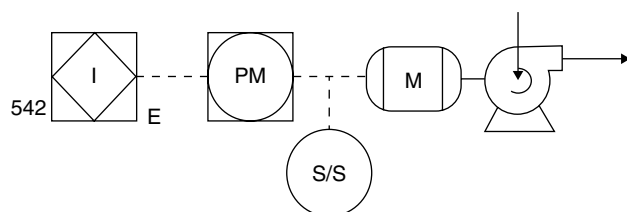


Figure 16.12 Electric motor as SIS final element.

16.8 Showing Safety Instrumented Functions on P&IDs

Up to now, we have learned how to show different SIS elements on P&IDs. Now we want to learn how to show a whole function. Let's have a look at the two examples below.

In the first example, when the level gets too high, the SIS action could be to close the inlet valve and open the outlet valve (Figure 16.13).

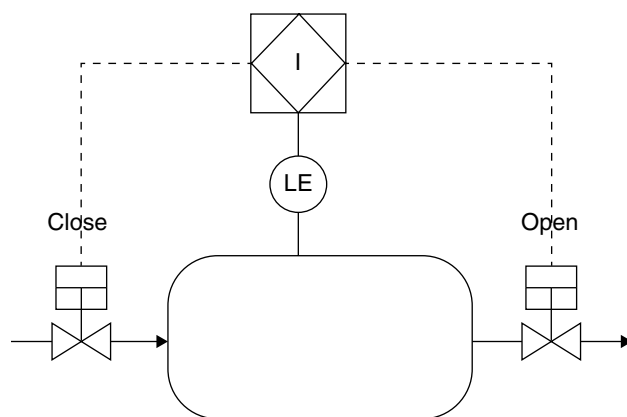


Figure 16.13 Example of a SIS.

In the second example, to protect a pump from cavitation, we can implement a safety feature to shut down the pump when the NPSHA is too low, or the pressure of the vessel is too low (Figure 16.14).

There are two issues regarding this type of representation of SIS functions: one is crowdedness, and the other is confusion in the interpretation of the SIS function.

Showing such simple cases on a P&ID is very easy, but more complicated SIS functions, when they involve "and/or" functions, are difficult to depict on a P&ID, and it makes the P&ID very crowded.

Regarding the second issue, you may ask: "in the top example, how do we know that the left switching valve should be closed and the right switching valve should be opened when the level reaches high-high?" This can be interpreted in the opposite way, i.e. "the left switching valve should be opened and the right switching valve should be closed when the level reaches high-high!" This example was very obvious, but sometimes it is not so easy to work out what effect the SIS action will have.

The solution to these two issues is to remove all the lines connecting different elements of SIS functions (decrease crowdedness) and to refer to another table for the interpretation of the SIS function (clarity of interpretation).

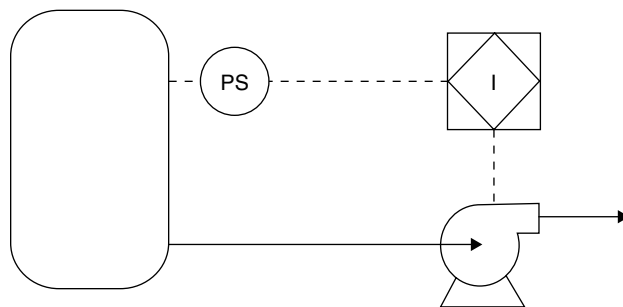


Figure 16.14 Example of a SIS.

We can refer to a “cause and effect table” (also known as a “shutdown key”), to see how the SIS operates. An example is shown in Figure 16.15.

A discussion on the cause and effect table is beyond the scope of this book.

By using this “trick,” the SIS function shown at the top of Figure 16.15 can instead be shown as depicted in the bottom of the figure.

The top arrangement in Figure 16.16 shows a level element, or level switch, going to a switching valve which can activate the pneumatic piston valve.

As a side note, both control valves and switching valves have air tubing and solenoid valves associated with them. We never show instrument air tubing around a control valve, but some people like to show air tubing and associated solenoid valves around switching valves. However, this tends to make the P&IDs very crowded.

Another way of depicting the SIS mechanism is shown in the bottom schematic of Figure 16.16. Instead of showing a connection between the level switch on the tank and the control valve, we can show an interlock tag (I) within a diamond that relates to both the level element and the

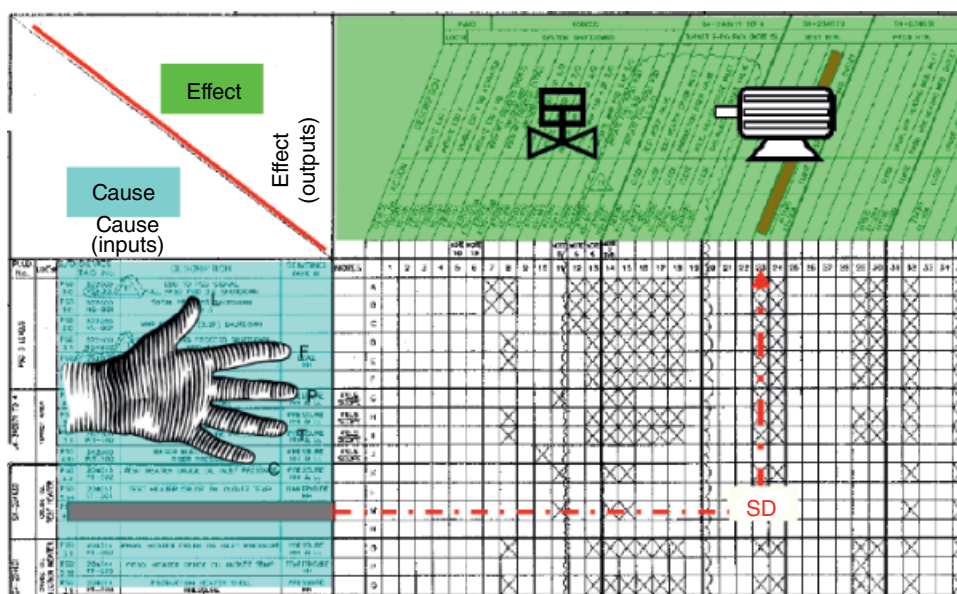


Figure 16.15 Shutdown key.

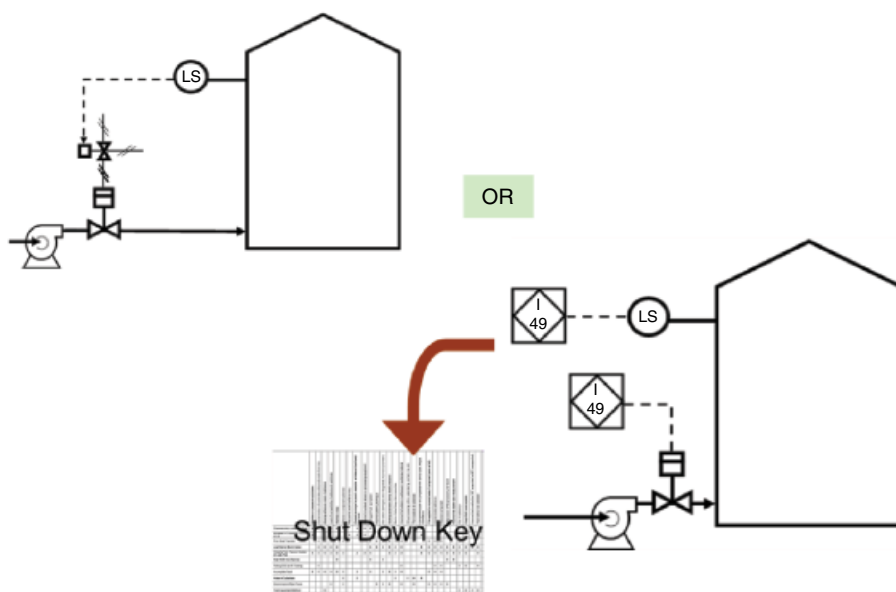


Figure 16.16 Simplified depiction of SIS functions on a P&ID.

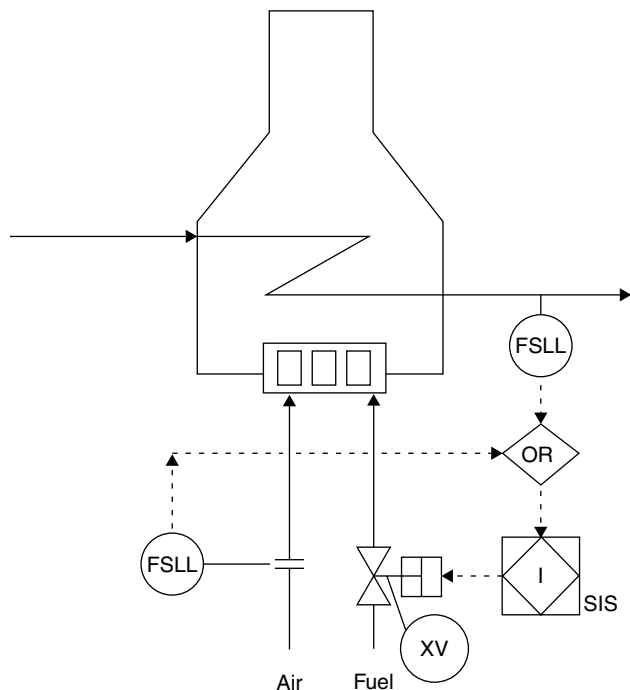


Figure 16.17 SIS symbology on a fired heater.

control valve. This takes the place of the solenoid valve and all the instrument tubing. If you are not sure how the interlock system works in this particular case, you can refer to the cause and effect diagram (shutdown key)

using the identification number 49, and this will tell you what action will be taken. In this case it says SD, which means shutdown of the electric motor on the pump.

Let's have a look at some other examples of SIS symbology.

Figure 16.17 shows the SIS control of a fired heater. It also shows the logic used for the safety function, through the use of an "or" function. This means that if we have low-low flow of product through the tubes or low-low flow of air to the heater, the SIS interlock will activate the pneumatic piston valve on the fuel line. This is a straightforward example, so you can imagine how crowded a P&ID would be if we opted to show all instrumentation lines and logic functions for the whole plant.

A variation of this symbology is shown in Figure 16.18.

Instead of showing interconnecting lines and logic functions, we can show the interlock symbols attached to the instruments. In order to see what the SIS control mechanism is, you would refer to number 214 in the shutdown key. Some people condense the symbology even further by combining the two diamonds of each instrument, as shown in Figure 16.19.

Figure 16.20 shows another example of SIS control symbology.

If we read the logic behind this function, it shows that if we have a low-low level in the tank on the left, or a high-high level in the tank on the right, the SIS will be activated. It does not show precisely what the SIS action will be, but an educated guess would be to shut down the

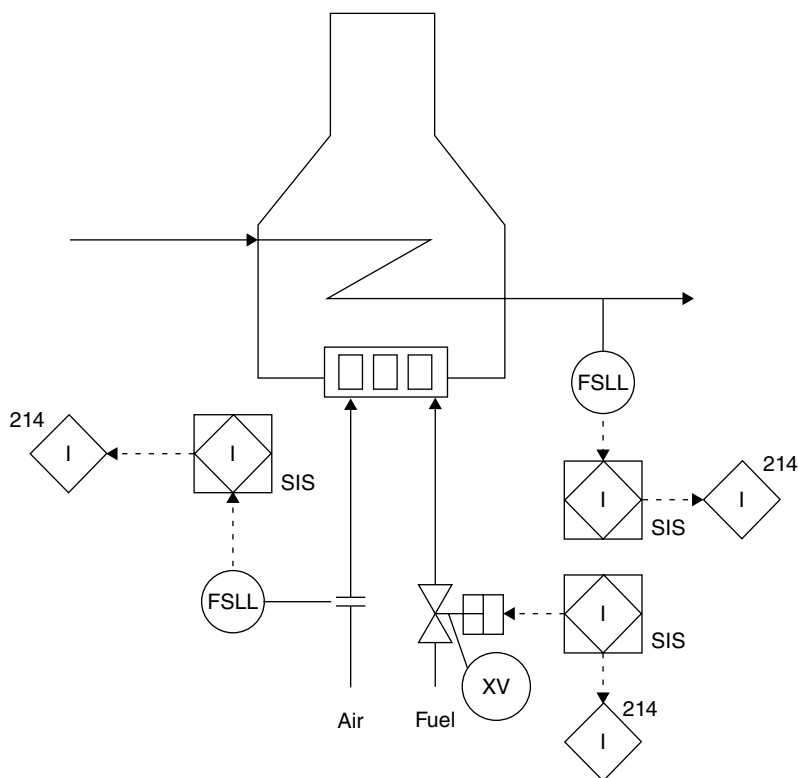


Figure 16.18 Variation of SIS symbology on a fired heater.

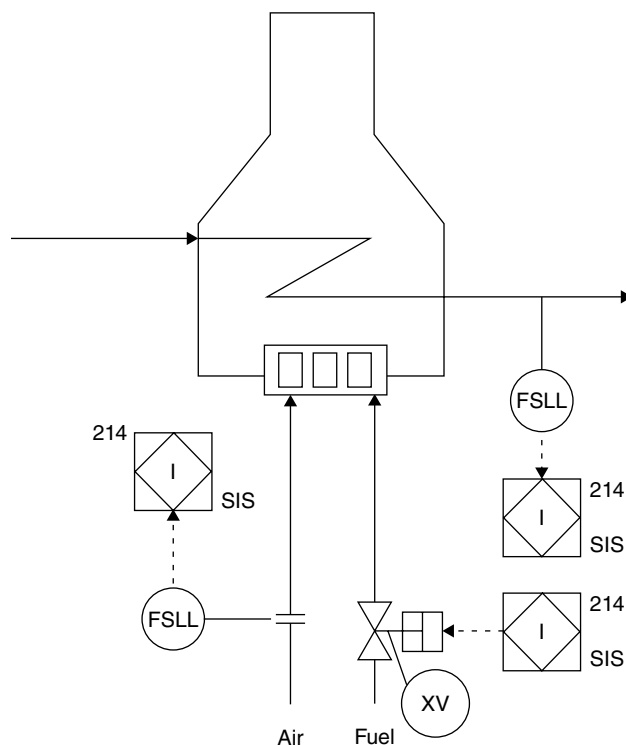


Figure 16.19 Variation of SIS symbolology on a fired heater.

pump. (But maybe it will start up the pump, who knows? The only way to check this is to refer to the cause and effect diagram.)

16.9 Discrete Control

“Discrete BPCS control” or simply “discrete control” is very similar to SIF actions, but it is not used for safety reasons. Discrete control is part of a BPCS action, but the final action is an on/off action of a switching valve

and/or a rotary machine. This control mechanism is activated for process, and not safety purposes.

If we can call a SIS a “safety interlock,” then the name of discrete control would be “process interlock.”

Discrete control is used for non-continuous units. A non-continuous unit could be a semi-continuous unit, such as filtration, or a purely intermittent unit (batch operation), such as the units in pharmaceutical plants.

For example, in an ion exchange treatment plant we need to regenerate the ion exchange medium from time to time. For this purpose we have to open a specific switching valve, and this is done for process and not safety reasons. Although the valves are switching valves, we do not tag them as XV or UV. We generally tag switching valves in discrete control as KV because the K indicates a time-based operation and this is usually the case in discrete operations.

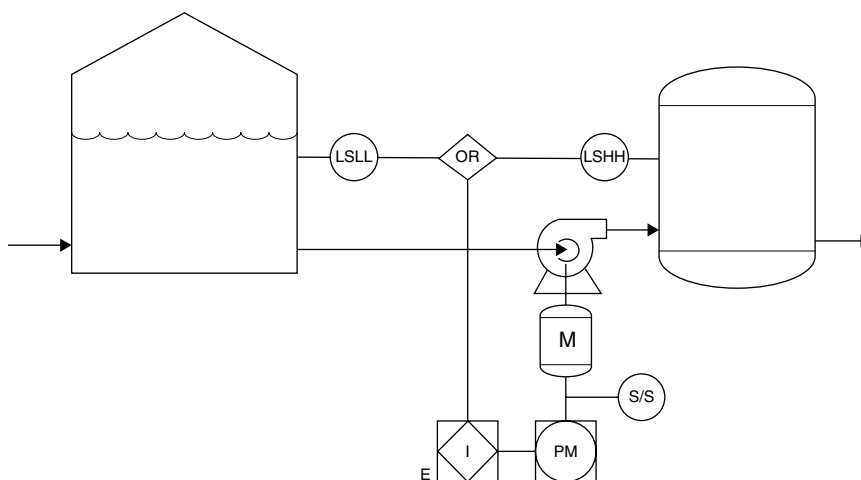
In the schematic in Figure 16.21, we have a switching valve, tagged KV 218. This is controlled by an interlock that can receive two signals, 435 and 213. We need to go to the shutdown key to see what kind of action is involved. Also, you can see that the interlock is marked as part of the BPCS and not the SIS, so we can recognize that this is a discrete control system.

As was mentioned, we may choose to show the switching valves as depicted in either of the schematics. The left-hand schematic in Figure 16.21 shows the brief depiction, while the right-hand schematic represents the more detailed arrangement.

On the right-hand side, we show the details of the actuator with its solenoid valve and instrument air supply. The variation on the left-hand side does not show this detail but in reality, you will see these instruments in the field.

The schematic in Figure 16.22 shows another example of discrete control. This could be an example of an intermittent operation (batch operation) to produce an

Figure 16.20 SIS control on two interconnected tanks.



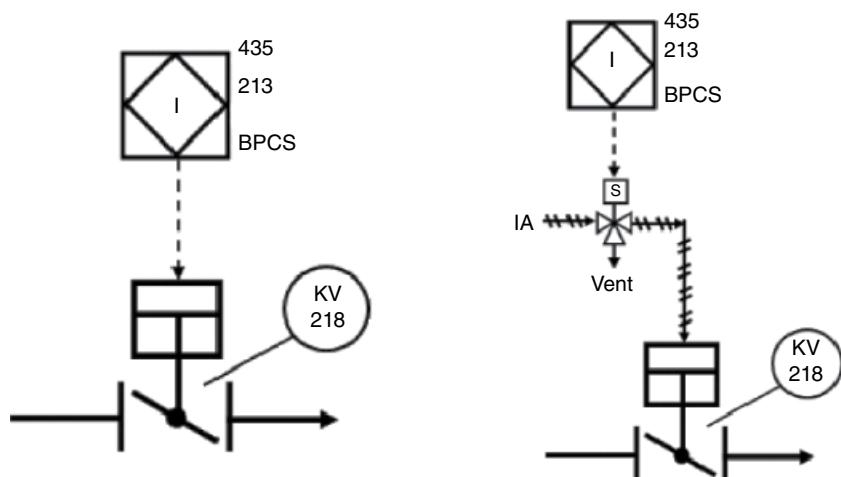


Figure 16.21 Discrete control variations.

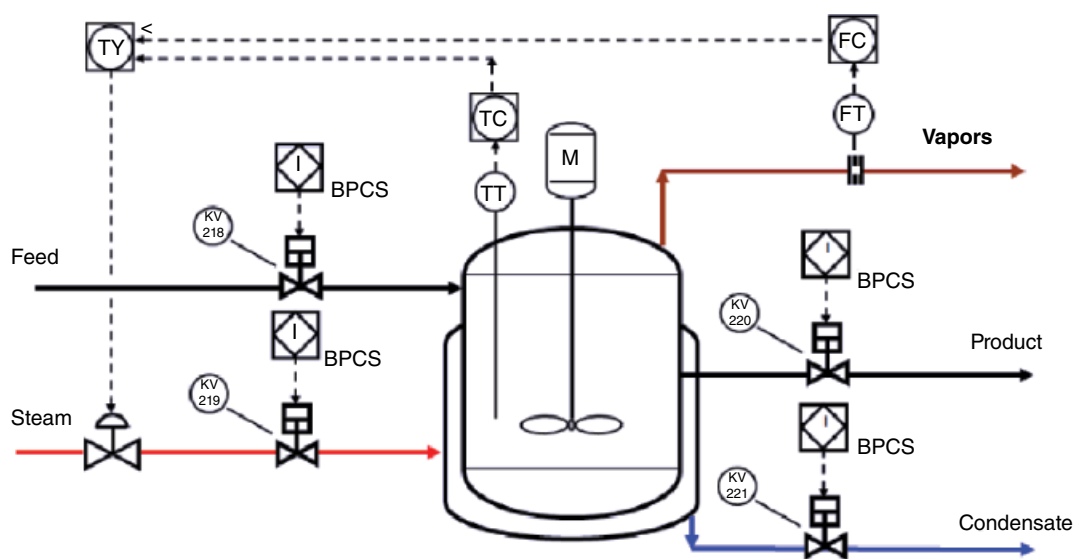


Figure 16.22 Discrete control in a batch system.

expensive product such as a pharmaceutical. Here you can see that besides the discrete control elements, which can be recognized by their “diamond” shape, there is one BPCS control too.

The BPCS control here continuously controls the flow rate of steam to the reactor jacket to maintain a specific temperature in the reactor.

We can recognize that this arrangement as a batch system because we have a time-based control valve, KV, on the feed line. For this reason, we need to install discrete control valves on the steam, condensate, and product lines. The fact that the jacketed vessel needs steam shows that the reaction is endothermic, which means it needs heat to proceed.

16.10 Alarm System

The objective of an alarm system is to warn operators. There are mainly two types of alarms.

One group is the ones triggered when the process parameters are exceeded and before a SIS is going to be activated but the process fluid is still inside of process units.

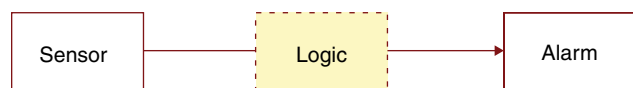
The second group is the alarms that work when process fluid is already discharged or released from the process unit, which means loss of containment happened.

Table 16.3 provides summary of features of two types of alarm systems.

We discuss these two types of alarm systems in Sections 16.10.1–16.10.4.

Table 16.3 Features of two types of alarm systems.

	Group 1	Group 2
Name	Alarm system	Fire and gas detection system (FGS)
Schedule of action	Before loss of containment	After loss of containment
	Point sensitive	Space sensitive
	Inside of units	Outside of units
Target parameters	Temperature, pressure, flow rate, level, composition and non-process parameters	Only fire and gas
Footprint on P&ID	Can be seen on every sheet of P&ID	Could be only several sheets dedicated to it as “FGS P&IDs”

**Figure 16.23** Anatomy of an alarm system.

Alarm management specifies all the activities related to adding an efficient alarm system in a plant. Alarm management requires a specific skill and some control engineers are specialist in this area. Here, again, we do not want to go to deep knowledge of alarm management.

16.10.1 Anatomy of Alarm Systems

Figure 16.23 shows the anatomy of an alarm system.

In an alarm system there is a sensor that sends a signal to a logic processor, and then to an alarm.

The primary element of an alarm system could be a sensor or switch. The sensor/switch could be linked to any of the process or non-process parameters such as level, flow, temperature, pressure or composition.

The logic could be a simple logic handled by a DCS or PLC. There are even cases that for a single parameter one alarm, e.g. at low level, is handled through a DCS but the low–low level of the same parameter is handled by a PLC system.

There are also cases that the primary element of an alarm system is connected to the final element through hard wires. These alarms are named hard alarms to be recognized from “soft alarms”, which are through a DCS or PLC.

The final element of an alarm system or alarm could be classified in different ways. Alarms can be audible or visual and can be in on monitor screen or a stand-alone type.

Table 16.4 shows different types of alarms.

Alarms can also be classified by their location, i.e. in the field or in the control room. When they are in the field, they are definitely stand-alone alarms. The field alarms are mainly audible type.

Table 16.4 Different types of alarms.

	Audible alarms	Visible alarm
Monitor screen alarm	Beeper	Flashing icons
Stand-alone alarm	Buzzer, horn, siren, bell	Flashing lamps, rotating lights, strobe lights, beacon

When they are in the control room they are, these days, on monitor screens. There could still be, in some cases, alarms in control rooms that are installed on a panel, in the form of stand-alone alarms.

16.10.2 Alarm Requirements

An alarm is an alert system to notify the operator of an “upset” event. The logical reply for an alarm is an action or a set of actions. The action could be done by SIS (automatic actions) or by plant operators.

There are two questions regarding alarms on P&IDs that needs to be answered: which parameter needs an alarm, and which level of parameter needs to be used for the alarm triggering point.

Deciding about parameters needing an alarm is not always easy.

The designer has to optimize the number of alarms in the system. Very few or too many alarms are dangerous for a plant. If the intent is very few alarms, some important alarms may be ignored, which is dangerous. Too many alarms may lead to a situation where the operator becomes insensitive to them and doesn’t deal with them with the required degree of urgency. This is again dangerous for the plant.

In some instances, if there are too many alarms in the control room, the operator becomes overloaded and some of alarms may be overlooked.

One rule of thumb is that if a SIS exists for a parameter, it definitely needs an alarm too. If a safety interlock

function is already added to a process parameter, it definitely it needs an alarm too. However, the reverse is not true in all cases. It means there could be cases that an alarm is added to a process parameter but the safety interlock function doesn't exist. These are the cases that an alarm is needed for the action of operators.

To investigate if an alarm is needed even where no SIS exists a sort of mini-HAZOP should be done.

It should be noted that if an event calls for an alarm it is generally wise to have a SIS on it too. If there is no SIS on it, it is probably not possible from technical viewpoint. The need for alarm can be checked based on probability and the consequence of the bad event. If the probability of a bad event is high and/or the consequence is high, it may need alarm, even in the absence of a SIS. When talking about consequence we need to think about safety, environment, equipment failure, and process upset. A severe equipment failure is the one that makes a whole piece of equipment out of order permanently and/or creates loss of containment.

To answer to the second question it should be said that an alarm could be placed in one level for a specific parameter or more than one level. When they are in more than one level, it means they have different levels of criticality.

The position of an alarm for single level alarm is shown in Chapter 5, Figure 5.13.

Some plant owners prefer to have two sets of alarms: a warning pre-alarm, and alarm. The pre-alarm is usually set to activate just before high or low levels and serves to direct the operator's attention to a potentially critical situation so that he can take some corrective action (Figure 16.24).

Depending on the type of alarm, the sensor could be set at the high, low, high-high and/or low-low positions. However, the high-high and low-low levels are generally "reserved" for SIS actions and it is more common to see alarm of high and/or low levels.

Some other plant owners use three level alarms; advisory, warning, and critical.

The different level of criticality can be recognized with different colors in the case of visible alarms or different pitches in the case of audible alarms.

The concept of different criticality levels of alarms is not only applicable for a single parameter but also for different parameters associated with different units in the plant. It means that all the alarms in a plant can be classified in two or three levels of criticality.

The different levels of criticality for alarms are not generally stated on P&IDs. They are primarily found in an alarm table.

16.10.3 Alarm System Symbolology

The primary instrument in an alarm system is the sensor that has a symbol similar to symbols shown in Chapter 14 for different sensors.

The logic for alarm system is run through PLC or a DCS. It means that an alarm can be incorporated into a BPCS or a SIS. The decision for handling an alarm through a BPCS or SIS depends on the availability and also criticality of an alarm.

On the other hand an alarm can "reside" in an alarm server as a stand-alone alarm or it could be part of controller. Generally speaking the more important alarms should reside in an alarm server as stand-alone alarm. Where an alarm is not stand-alone, it could even be shown as part of an "indicator." However, in such cases the indicator cannot work as alarm logic solely. There should be logic that has an alarm set point and triggers it.

Therefore, there could be four different symbols for the logic of alarms (Table 16.5).

As can be seen in the table, when the alarm is stand-alone, the letter A – standing for alarm – clearly indicates that the temperature parameter is activated at the high level setting. When an alarm is not stand-alone, on the side of the TI balloon, there is H to indicate that the alarm is activated at the high level setting. However some companies prefer to use TAH beside the balloon to clearly show the existence of an alarm.

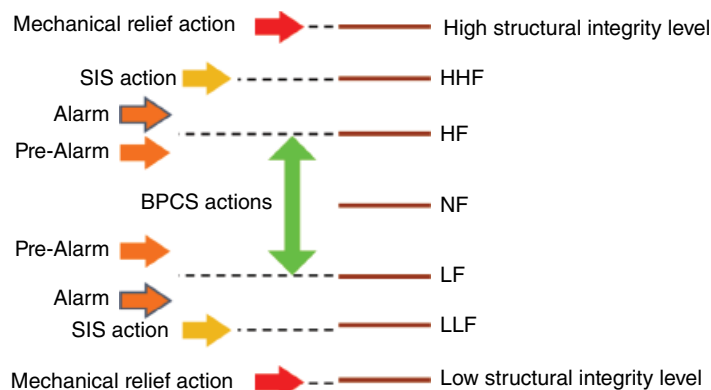


Figure 16.24 Pre-alarm and alarm positions on a parameter level system.

Table 16.5 P&ID symbol for the final element of alarm logic.

	Alarm through BPCS	Alarm through SIS
Stand-alone alarm		
Alarm as part of other parameter monitoring instruments		
	Or: 	Or:

The last element of an alarm system is an “annunciator.” An annunciator is catch-all word for a visible and/or audible alarm final element.

The annunciators could be recognized through a monitor screen in control room, or could be in the field.

It can be arguably said that alarms are placed in a control room by default to warn the control room operators. However, there are cases that a field alarm is needed in addition to the control room alarm. Those are the cases that if the required action is not very complicated and/or the danger is so high that it should warn the field operators to move away from the area (evacuation).

Table 16.6 shows different symbols for alarms in the field and in the control room –through a monitor screen – in the form of audible or visible.

When an alarm is in the field, it is physical device and it is a good idea to show if it is audible or visible.

Table 16.6 P&ID symbol for alarm logic.

	Field alarm	Monitor screen alarm
Visible alarm		
Audible alarm		

When an alarm is through a monitor screen, its type – visible or audible – is not very important and the general symbol for an annunciator could be adequate.

In either a field alarm or a monitor screen alarm, the final element of an alarm can be disregarded. This means that in some P&IDs, the final element of an alarm is not shown and only the alarm logic symbol is left, or they merge it with a symbol of alarm logic.

Therefore there are plenty of different ways are available to show an alarm “loop.” Figure 16.25 shows this concept. There is an alarm system on high level of temperature that the final element is an alarm on monitor screen of HMI. Figure 16.25(a) shows a complete representation of this system. In Figure 16.25(b) the symbol of logic element is deleted for simplicity. In Figure 16.25(c) the symbol of final element is deleted. In Figure 16.25(d) a field alarm is added, which is an audible alarm. Figure 16.25(e) is similar to Figure 16.25(d) except for the type of field alarm, which is changed to visible alarm.

In the examples of Figure 15.25 all alarms are functioning out of the DCS and the alarm resides in the alarm server.

Figure 16.26 shows several examples of alarms on P&IDs.

16.10.4 Concept of “Common Alarm”

There are some cases that several alarm systems merged together. In such arrangements, those several alarm loops keep their own primary elements and logic, but there is a single final element or single annunciator. Such alarms are named “common alarm.” In the common alarm systems, when the alarm starts to warn, the operator cannot recognize the exact point of fault or bad event. He needs further investigation to find the problem.

Common alarms are very popular for compact items with the chance of failure in their sub-systems. One famous example of a common alarm is generally in electric motors (Figure 16.27).

16.11 Fire and Gas Detection System (FGS)

This system is to monitor the existence of fire and non-innocent gases in the spaces of process plants.

In the majority of cases the existence of the above-mentioned hazardous elements trigger the FGS, which generally is alarming. There are, however, some cases where the FGS triggers an action, or set of actions. For example FGS may trigger a firefighting system.

The sensors for fire could be anything like a flame sensor or smoke sensor.

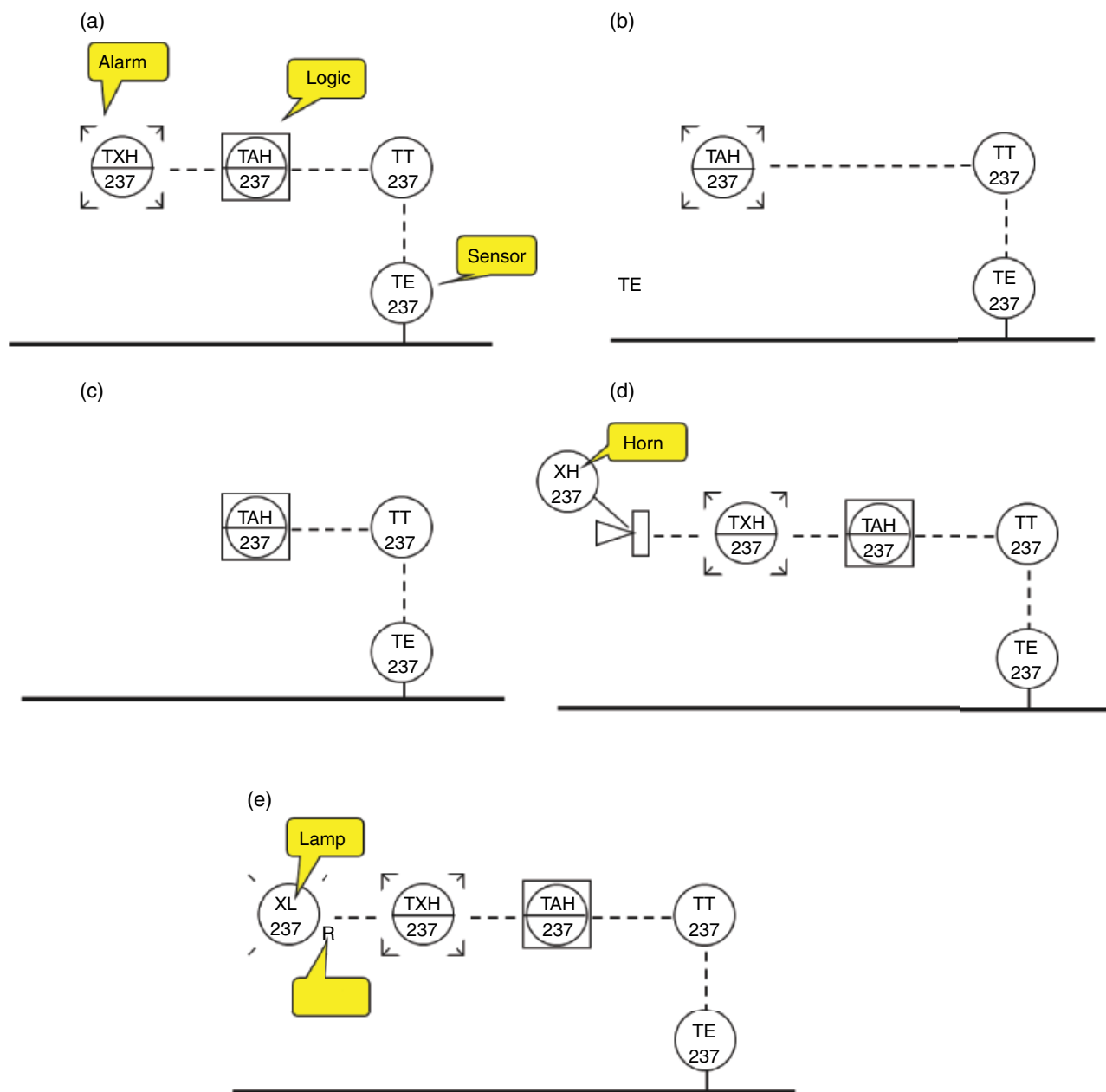


Figure 16.25 Different ways of showing one alarm system on a P&ID.

The type of gas detection system depends on the types of gas to detect. In an FGS any aggressive gas can be decided to be measured. The gas of interest could be a flammable gas, toxic gas, or even inert but suffocating gases like nitrogen gas.

The anatomy of the FGS is very similar to that of a SIS, except that the final element is not something that interferes with the process (like a switching valve); it is only an alarm.

At the P&ID level, there are two aspects of FGS: its signal handling system (control system) and the location of

sensors and alarms. The signal handling system can be shown in auxiliary P&IDs. Figure 16.28 shows a typical FGS control system that can appear on an auxiliary P&ID.

Figure 16.28 shows a flammable gas detection system that:

- Has a sensor ("AE"), transmitter/indicator in field ("AIT"), indicator in control room through PLC system ("AI" in diamond).
- Alarms when the flammable gas concentration is 20% of the LEL low explosion level); alarm-analyte is high level.

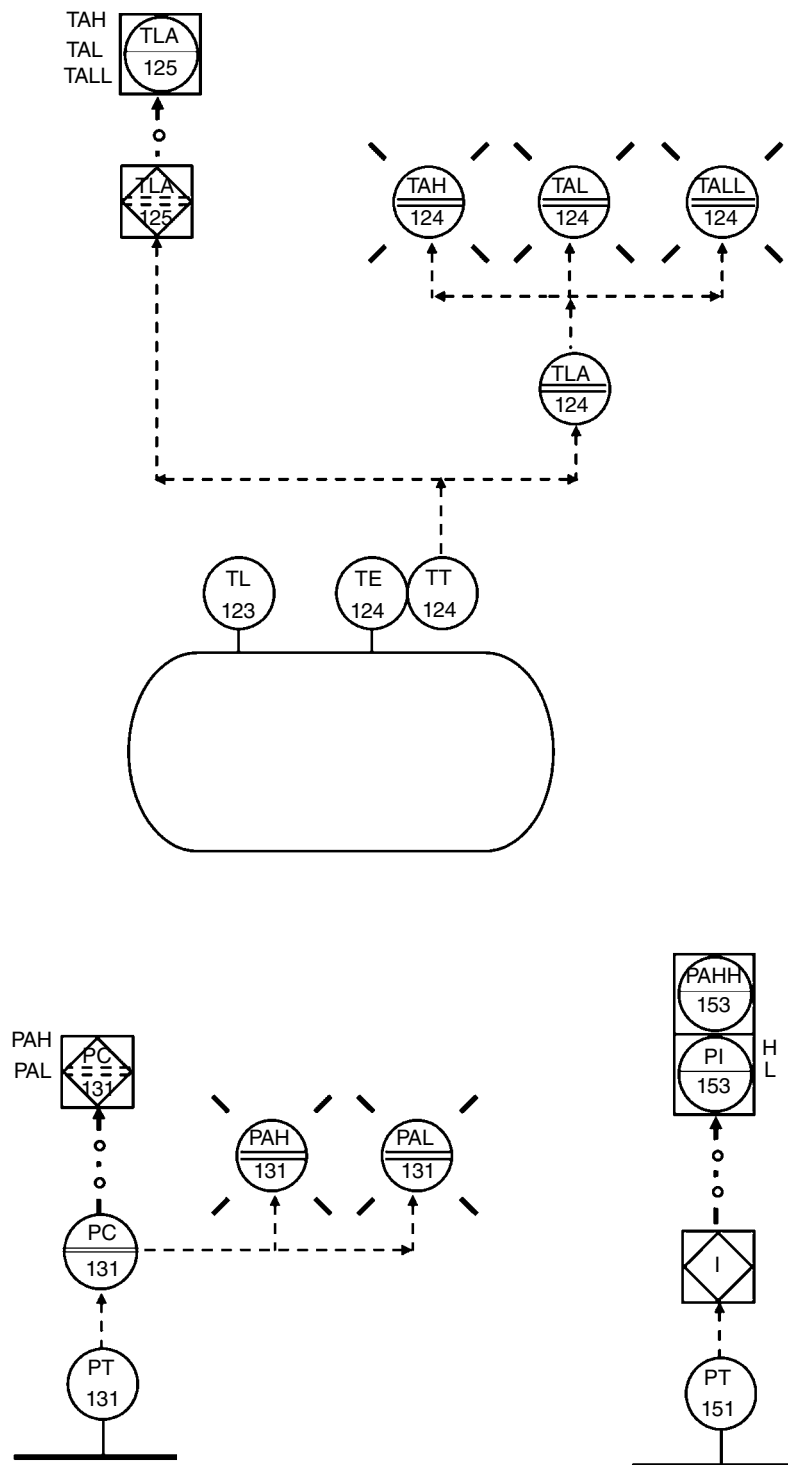


Figure 16.26 Some example of alarm systems on a P&ID.

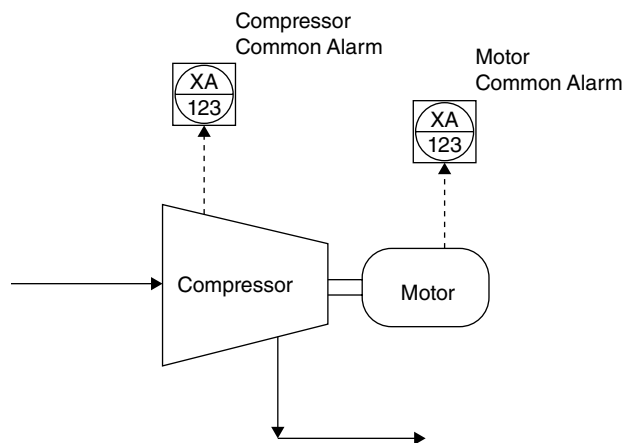


Figure 16.27 Common alarm on P&ID.

- Alarms when the flammable gas concentration is 40% of LEL; Alarm-analyte is high-high level.

The location of sensors doesn't need P&IDs but they need to be shown on plot plant (Figure 16.29).

16.11.1 Manual Alarm

Manual alarms are generally in the form of audible ones like a horn or beacon. These are for operators to warn others about an immediate danger.

There are two aspects of manual alarms that should be engineered; their locations and their hardware detail.

Their location can be shown on a drawing based on a plot plan.

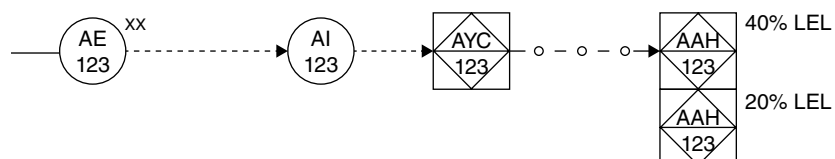


Figure 16.28 Flammable gas detection system.

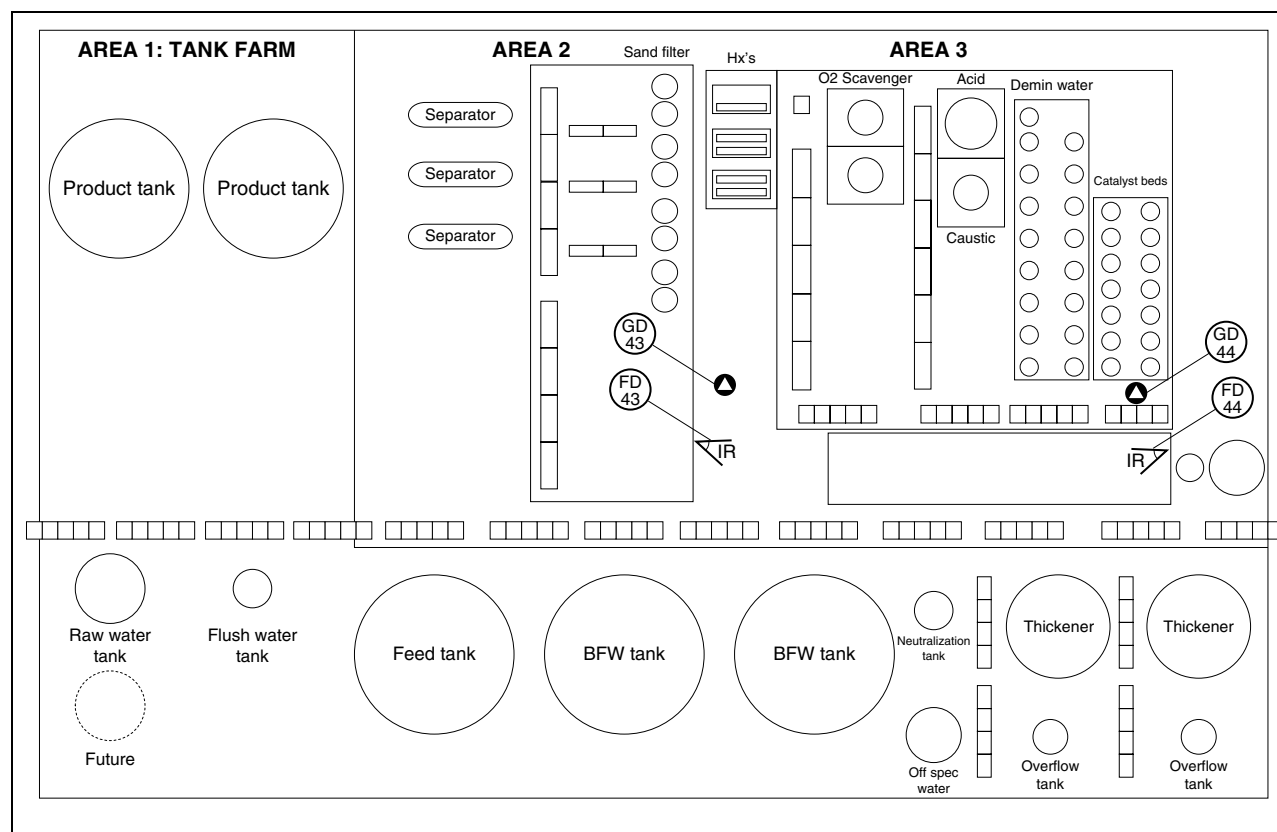


Figure 16.29 Sensor locations of a FGS.

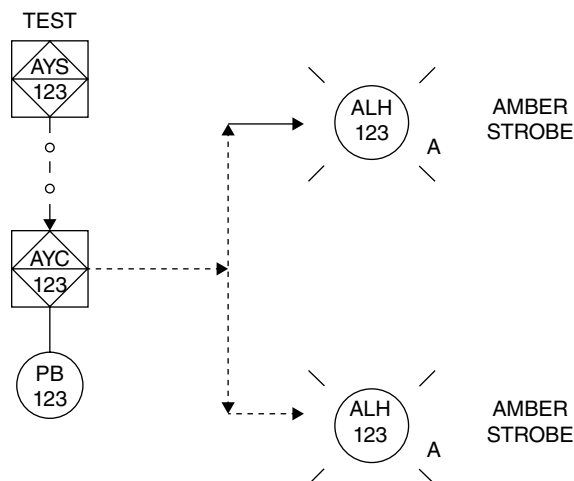


Figure 16.30 Manual alarm system.

The detail of hardware can be shown on the P&ID as an auxiliary P&ID (Figure 16.30).

16.12 Electric Motor Control

The discussion of electric motor control has been moved from Chapter 15 to here, so that we can discuss both BPCS and SIS controls for electric motors at the same time.

Different companies have different opinions regarding how to show electric motor control. Some companies have a very simplistic approach and show only one box for electric motor control, whereas others may prefer to show a more detailed version of it.

We will start with a very simple depiction of electric motor control and move on to a very detailed one. You can think of the schematics here as merely an example of how to show electric motor control. Your company may decide to use a mixture of the symbology shown here to depict electric motor control.

16.12.1 Simple Motor Control

Figure 16.31 shows several representations of motor control in different companies. Some companies don't show electric motors at all (top left). Some other companies show the motor without any control system on it. This schematic shows examples of how different companies may illustrate motor control. PM stands for pump motor; if the motor were connected to a compressor, it would show CM.

16.12.2 The Focal Element of Motor Control: MCC

In a more detailed way of showing electric motor control, we may show MCC, or motor control center. I usually

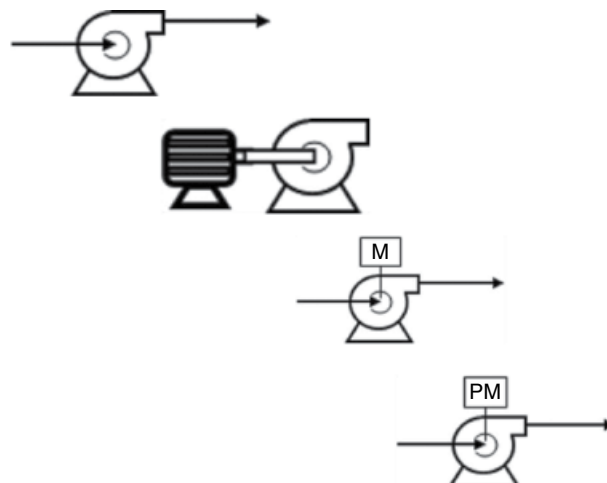


Figure 16.31 Illustration of motor control.

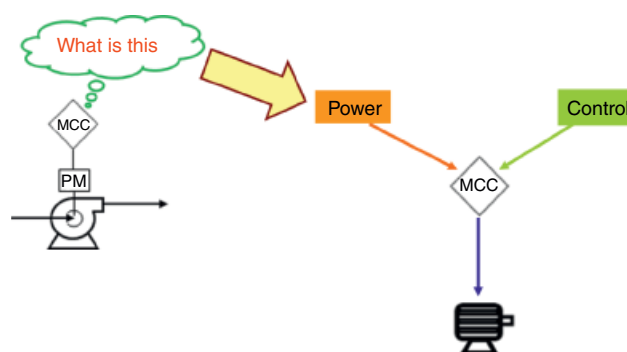


Figure 16.32 Motor control center.

describe an MCC, for people who are not versed in electrical engineering, as a “meeting room” for power signals and control signals! (Figure 16.32).

Within the MCC, we must have a power signal to drive the motor. We also have a control signal coming in that will tell the motor to accelerate, decelerate, start up or shut down.

16.12.3 All About Relationships with Electric Motors

All relationships between plant/plant operators and an electric motor can be summarized under three categories: ordering motor, reporting by the motor, and principal arrangement for inspection and repair.

In category 1, the motor receives orders or “command” signals from the plant. A command going to the motor could be:

- A process command. This could be a regulatory command (BPCS), for instance from a VFD to increase or decrease the RPM of the motor. It could also be an interlock, either a process or safety interlock (SIS).

The type of orders by SIS could be start-up and shutdown of the electric motor.

- A manual command. This could be a command that comes in from the operator in the field or control room.

“Command” signals are the orders that are sent to a motor, or more correctly, to the MCC of a motor. C could be used as the representing letter for these types of functions in P&ID symbols.

“Command” signals are always available around a motor because they are the arrangement to make the plant and/or operator to control a motor.

Category 2 is the reports provided by the motor. “Response signals” are the reports that are generated by the motor, or more correctly, by the MCC of a motor.

There are mainly two types of signals: the signals that report if the motor satisfied a “command” and “responses” that report parameters on the “health” of the motor, which are running reports and trouble reports.

An example of command report signals is the signal is sent from the motor if it turned off after receiving a command for turning off or not.

An example of a running report signal is when a motor reports the total hours that it is working. This is especially important for motors connected to parallel pumps; they need to work roughly the same number of hours each to ensure their optimum health.

The other example of a trouble report signal is a “common trouble alarm.” This signal is very common to see and it is an alarm by the motor that warns the operator of some type of problem inside the motor.

“Response signals” are the “motor’s talking” that are sent by a motor, or more correctly, by the MCC of a motor. S could be used as the representing letter for these types of functions in P&ID symbols. A signal S could activate an indicator, a lamp or an alarm on the control panel.

“Response signals” are not always available. A designer may put them around a motor if it is critical to know the condition of the motor.




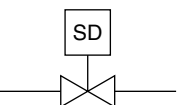





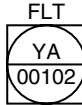

The main element of category 3 is HOA switch. The principal arrangement for inspection and repair is started with an HOA switch. However, there are some other switches around an HOA switch that need to work together to be able to perform a complete inspection and/or maintenance.

In Sections 16.12.4–16.12.6 the P&ID representation of three categories of electric motor functions are discussed.

16.12.4 P&ID Representation of Commands and Responses

As it was stated there are two types of signals. The signals that are generated in “reaction” to a command and the signals to report the health of a motor. These two types of signals are shown in Table 16.7.

Table 16.7 Symbols of commands and responses.

	Meaning	Representation
Commands 	Do it!	Automatic:
		Regulatory: 
		Interlock: 
		And in a combination form could be:
		
		Manual:
		In field: 
		In control room: 
		SS command
		or: 
Responses 	I did it sir! Here is the proof!	They could be in any of three types of indicators, alarms, or lamps.
		Indicator example: 
		(I at the end of tag)
		Alarm example: 
		(A at the end of tag)
		Lamp example: 


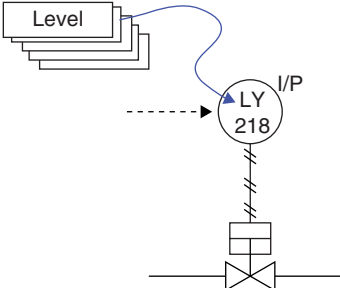

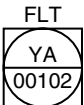

There are at least two issues regarding showing motor control in P&IDs.

The first one is that there are plenty of parameters involved that are not defined by the ISA. The solution is using non-specific letters from ISA, like M, N, or Y and then explaining them right beside the balloon. Usage of Y, however, is very common because it refers to any event or state.

Several examples are shown in Table 16.8.

The second issue is that each motor so many control items around it on P&IDs that sometimes it is not easy

Table 16.8 Examples of non-specific parameters.

P&ID symbol	Meaning of non-specific parameter
 SS COMMAND	X represent start-up and shutdown (SS)
	Y represent different parameter to control
	Y means operation. K means time. Q is total Then YKQ is "total operation duration."
 FLT	Y represents "fault" (common fault)
 STATUS	Y represents "status"

to show them all around a small motor symbol. The solution is the summarization of symbols.

Table 16.9 shows one example of this.

The problem with this type of "compaction" is it is not very easy to interpret.

16.12.5 P&ID Representation of Principal Arrangement for Inspection and Repair

"HOA switch" stands for a hand-off-auto switch, as shown in Figure 16.33.

An HOA switch is always used for rotating machinery. On a P&ID, we indicate an HOA switch with HS inside a circle and H/O/A outside it. This is how the switch works:

- Off. In the "off" position, the motor is totally switched off. No one in the control room can switch it on and nobody in the field can turn it on by using a local switch at the pump.
- Hand. When the switch is in this position, it means that the motor is disabled remotely but can be locally enabled. This means that you can turn it on in the field but not from the control room. When maintenance personnel are working on the motor, they switch the HOA to the "hand" position so that no one in the control room can switch it on by mistake. We said that when the HOA switch is on H it can be turned on or off from the field, so we need to show the field pushbutton for this action. Besides the symbol for the HOA switch on the P&ID, we have to show HS and S/S, which indicate a hand switch for start up or shut down.
- Auto. When the switch is on "auto," the motor is remotely enabled but is locally disabled. Remote activation means that it can be operated by pushbutton in the control room. Figure 16.34 shows two ways of tagging these control room push-buttons on a P&ID. The bottom tag has the letters PM, which means it is connected to a pump motor. It gives the sequence number 115. It has a divider, which indicates that it is situated in the control room, and it has a square around the tag circle, which shows that it is visible through the DCS. The other symbology shows a tag with XCR, the sequence number, and S/S, which indicates that it is a stop/start button.

In addition, the control room personnel should be aware of whether the HOA switch in the field is on H or O or A. There is also another indicator in the control room to show the status of the HOA switch through the DCS. There are two ways of showing this on the P&ID: one to indicate whether the switch is in the remote or

Table 16.9 Examples motor control.





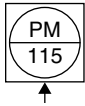
Expanded representation	Condensed representation
<div style="display: flex; justify-content: space-around; align-items: flex-start;"> <div style="text-align: center;"> S/S COMMAND STATUS </div> <div style="text-align: center;"> RUN STATUS </div> <div style="text-align: center;"> COMMON TROUBLE ALARM </div> <div style="text-align: center;"> L/R STATUS </div> </div> <div style="display: flex; justify-content: space-around; align-items: center;">     </div>	 S/S COMMAND RUN STATUS COMMON TROUBLE ALARM L/R STATUS STOP

Table 16.10 Signals in PBCSs and SISs.

	BPCS: Speed change	SIS: On/Off
Command		
Signal		

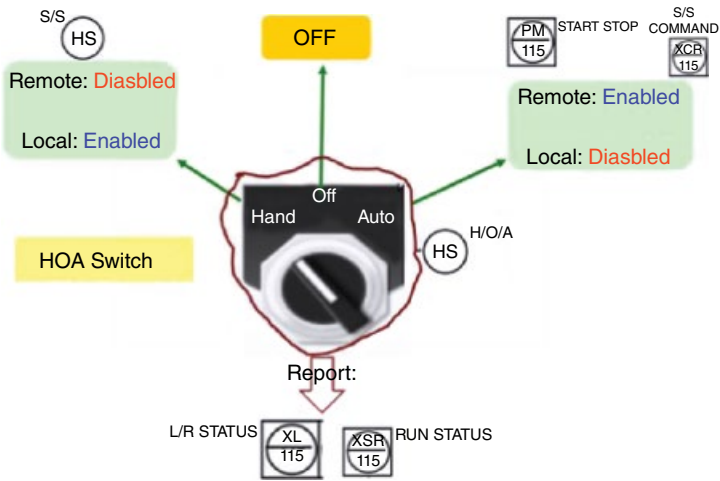


Figure 16.33 HOA switch.

Command			
Signal			

Figure 16.34 HOA symbology.

local mode (L/R status), and another to indicate the run status.

16.12.6 Examples

Figure 16.34 summarizes the different symbols that can be added to the MCC of a motor, based on an HOA switch.

Table 16.8 shows examples of P&ID representation for two types of signal in PBCS and SIS.

For example, in Table 16.8, you see in the BPCS that there is a command to adjust the RPM of the motor to a specific value. This command comes through an SC, or “speed control system” (some companies use the acronym VFD instead). When the motor performs this command, it generates a signal, SI, to show what the RPM of the motor is after performing the SC command. SI stands for speed indicator.

In the same Table 16.8 a SIS action can be seen too. The SIS function asks a motor to shut down through an interlock shown in a diamond. In return, the motor generates a “run status” signal to show what its running status is – running or not running – after the command.

“Health” reports by motors were discussed in the previous section. In Figure 16.35 you can see its examples.

In Figure 16.35(a) there is a report signal for a command signal of turning on or turning off of the motor. This report shows the running status of the motor.

Figures 16.35(b) and (c) show report signals related to the health of the motor. Figure 16.35(b) reports total hours that the motor is running and Figure 16.35(c) is only an alarm for every little failure of the motor, or common trouble alarm.

Now let’s go through a complete example of electro-motor control in Figure 16.36.

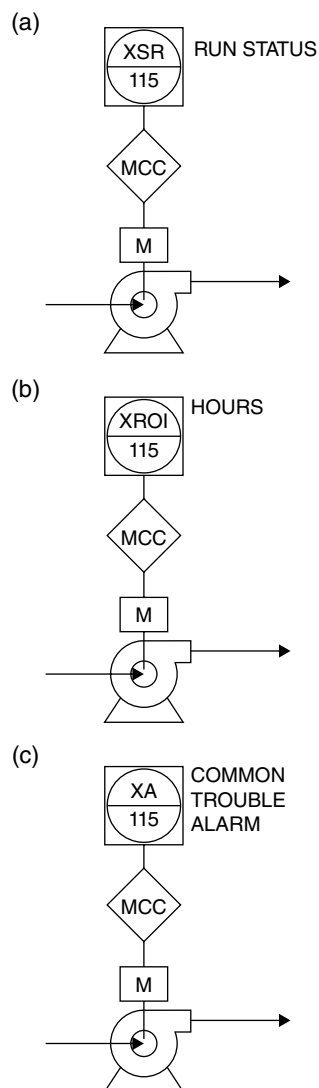
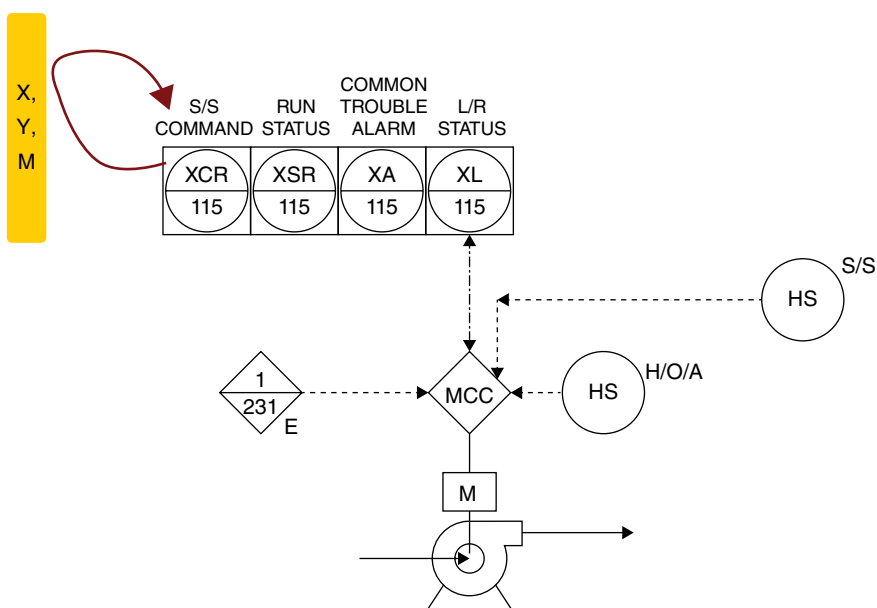


Figure 16.35 Reporting by motor.

Figure 16.36 Example of electromotor control.



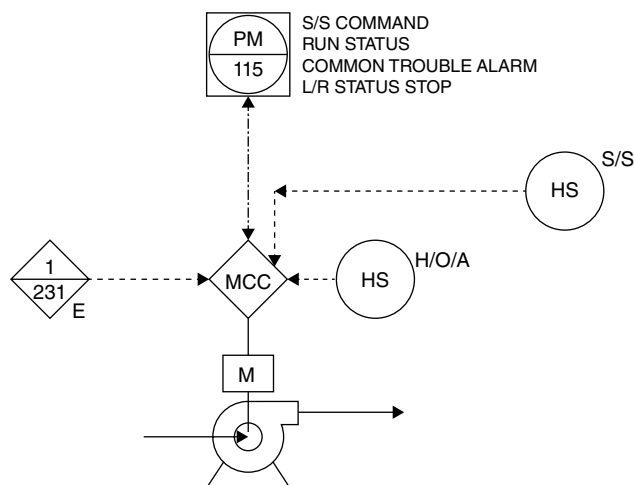


Figure 16.37 Example of electromotor control.

You have an MCC for the pump motor. Here we have already shown “MCC” and “HOA.”

If the HOA switch is set to “hand,” it means that control is remotely disabled and the schematic shows the hand switch, HS, for startup or shutdown (S/S). When the

HOA switch is set to “auto,” a signal will be coming in from the control room and the remote switch, HS, is disabled.

As you can see from the tags in the boxes, which have dividers through them, there are a number of indicators situated in the control room that show the status of the MCC. We have one for the L/R (local/remote) status, and others for common trouble alarm, run status and the start up/shut down commands. So, in effect, we have two-way communication between the control room and the MCC.

It is important to recognize that using an X (or sometimes Y or M) at the beginning of an acronym for motor control indicates a “non-specified parameter”; however, this parameter is immediately specified above or beside the balloon. For example, XCR-115 is a command from “X,” which is an “S/S command.”

There is also an interlock, I 231, on the motor control center. This signal comes from the SIS and we would have to look at number 231 on the cause and effect diagram to see what action would be taken in an emergency.

Figure 16.37 shows the same example as Figure 16.37 but in a more condensed form.

Part IV

Utilities

Utilities are auxiliaries of equipment and instruments and are required for their functioning.

As utilities may be needed by a large number of items in a plant, as an economic decision, they are mainly based on air and water. Ambient air free and water is abundant.

However there plenty of cases that should deviate from the basic air and water as the utility of choice.

These are mainly fluids but one important exception is electricity.

The fluids are used as utility to transfer material or to transfer energy.

The examples of utilities for the goal of material transfer are potable water, plant water, etc.

The examples of utilities for the goal of energy transfer are instrument air, cooling water, etc.

17

Utilities

17.1 Utility System Components

Utilities need to be generated in the plant, or purchased from another plant or company. For example, electricity can be generated in a small power plant inside an oil refinery, or it can be purchased from another external power plant. Steam is usually generated in a plant by a boiler or a steam generator.

To provide suitable utilities for each piece of equipment, the next step is to have a utility network. A utility network transfers the required utilities from the utility generator to the equipment. For example, if a piece of equipment needs steam, then steam needs to be routed from the boiler to that equipment. Because we may have multiple steam users in a plant, instead of having one pipe that is bringing steam from the boiler to equipment, we usually have a network of pipes that routes this steam from the boiler to multiple steam users. Therefore, in each plant, besides a string of equipment, we also have utility generation units and a utility distribution network.

Sometimes, after a utility is used it will convert to another type of utility, which needs to be recycled back to utility generation units. For example, after steam is “used,” it will be converted to condensate, and we don’t want to lose this by discarding it into the drain system. Plants like to save some money so we like to recycle this condensate to the boiler to convert it back into steam. So, for the majority of utilities, besides a utility distribution network, we also need a “utility collection network” for the “used” utility (Figure 17.1).

For example, when we have cooling water as a cooling medium utility for a system we generally call it “CWS,” which means “cooling water supply”; after this cooling water has been used in a system it is no longer cold, and this warm cooling water needs to be collected and returned to the system. We name this stream “CWR,” or “cooling water return.” Therefore, seeing – S and – R for heat transfer media is very common; these letters refer to supply stream and return stream, respectively.

Table 17.1 shows the pairing nature of some utility streams.

Not all the utilities have a collection network; for example, fuel gas could be used as a utility in a plant, but this utility generally doesn’t have a “collection system” per se, because the fuel gas will be converted to flue gas and released into the atmosphere after “use.”

17.2 Developing P&IDs for Utility Systems

To develop P&IDs for utility systems different steps should be taken. They are:

- 1) Identifying the utility users
- 2) Deciding on the utility network topology
- 3) Designing the detail of utility network
- 4) Placing priority on utility users
- 5) Connecting each user to the utility network
- 6) Connecting the distribution network to the collection network, if any collection network is available.

These steps are explained below.

17.2.1 Identifying the Utility Users

There are plenty of utility users in process plants. There are no universal guidelines that specify the utility users. This is the reason it happens that a P&ID development engineer misses a utility user until late in the project.

Table 17.2 shows some typical utility users in process plants.

17.2.2 Utility Distribution and Collection Network Topologies

Utilities can be distributed in a plant through one of these arrangements: a tree distribution network (manifold distribution) or a loop distribution (grid or mesh distribution).

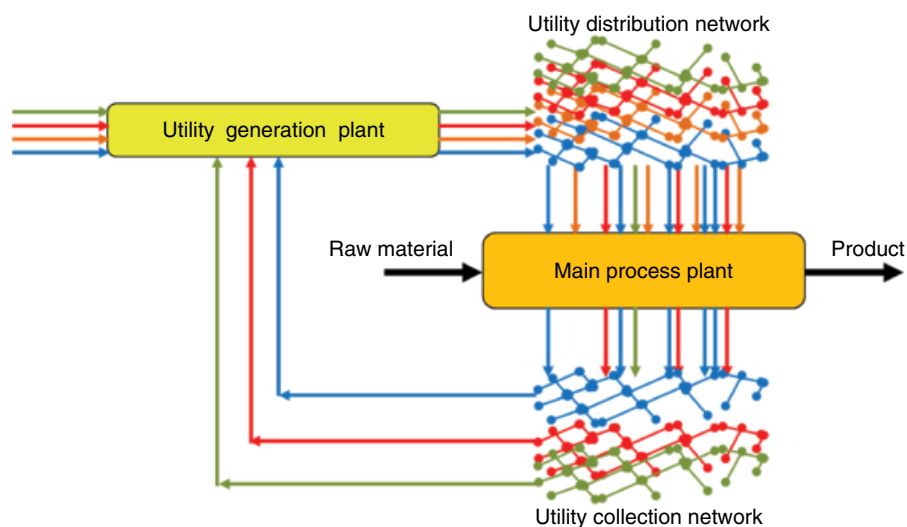


Figure 17.1 Utility cycle of a plant.

Table 17.1 Various types of utilities and their collection networks.

Utility distribution network	Utility collection network
Steam	Condensate
Cooling water (cooling water supply: CWS)	Cooling water return: CWR (which is warmed cooling water)
Hot glycol (hot glycol supply: HGS)	Hot glycol return: HGR (which is cooled hot glycol)
Cold glycol (cold glycol supply: CGS)	Cold glycol return: CGR (which is warmed cold glycol)
Electricity	–
Blanket gas	Vapor to the vapor recovery unit (VRU) system

Examples of these two networks are shown on Figure 17.2 for four utility users.

In both examples, the relative locations of users are the same to show how tree or loop distribution can be used for similar relative user positions.

Table 17.3 lists the features of each type of pipe networking.

In a loop distribution network there is more freedom to take care of more important users. The more critical user note should be supplied by more pipes. This concept is shown in Figure 17.3.

It is important to realize that the flow rate and the pressure in the utility network are always fluctuating. This is especially true for utility networks that have more intermittent users.

One problem of loop distribution is that it is the cause of decreased accessibility. It is obvious that loop distribution needs more pipes, which generally need to cross through different areas of the plant.

Table 17.2 Some typical utility users in process plants.

Utility types	Typical Users
Instrument air (IA)	Control valve, some sensors, some process analyzers
Utility air (UA) or plant air (PA)	Air-operated pumps, air cushions in silos
Utility water (UW) or plant water (PW)	For cleaning purposes
Potable water	Safety showers, eye washers
Heating media	For heaters
Cooling media	For coolers
Steam	For stripping: strippers, distillation towers, de-aerators For heating: heaters
Cooling ambient air	For combustion: burners For cooling: air coolers
Blanket gas	Blanketed containers
Fuel gas, fuel oil	Burners
Vapor collection network	Blanketed containers
Emergency vapor/gas release collection network	Safety devices, emergency blow down systems
Surface drainage collection network	Indoor areas, outdoor area for process pad
Fire water	Everywhere, but the insurance company needs to be consulted
Inert gas	Equipment deal with flammable gases
Condensate collection network	Equipment use steam for heating purposes
Electricity	Fluid movers, mixers, motor-operated valves

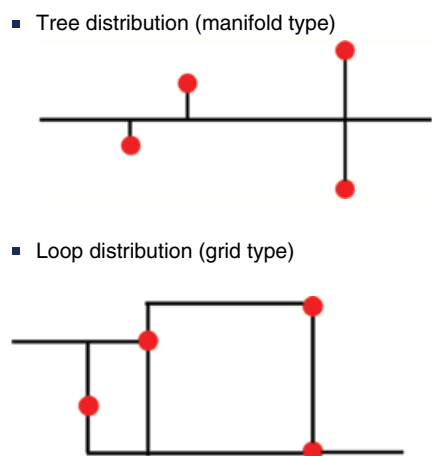


Figure 17.2 Utility distribution networks.

If the pipes are above ground, this quantity of pipes decreases the accessibility to different locations and equipment. One solution is routing pipes in a loop distribution system over the pipe rack or laying them down underground. These solutions, however, don't resolve the problem completely and loop distribution is still considered as a less favorable option.

One deviation of a loop distribution network is a tree distribution with recirculating fluid. One famous example of this distribution system is an ultra-pure water distribution system in pharmaceutical and microchip manufacturing plants (Figure 17.4).

One important example distribution of fuel oil amongst several burners of a fired heater. As the fuel should be distributed as evenly as possible, the distribution system is loop type.

17.2.3 Designing the Detail of a Utility Network

As the distribution and collection networks are generally not designed during the design phase of projects, it

should be at least cursorily checked during the P&ID development.

The design of distribution and collecting networks is very tedious and complicated if it is intended to be very accurate. However, such accuracy is not generally necessary for utility distribution and collecting networks. The reason is that such networks have plenty of parameters beyond the control of the designer. The very accurate design of distribution and collecting systems are generally done for the design of manifolds.

However, a list of qualitative rules of thumb can be provided for the check during the P&ID development phase of project:

- 1) The importance of distribution networks tends to be more critical than collecting networks.
- 2) In distribution and collecting networks the arrangement should be symmetric as much as possible.
- 3) To have more symmetric arrangement, use sub-headers in appropriate locations and add branches on them, preferably evenly.
- 4) Based on a continuity equation the total area of branches on a header or sub-header should be equal to the area of header or sub-header of interest. However practically we try to decrease the ratio of "total branch area to header area" to less than 0.9 and in large systems to 0.6 or even 0.5.
- 5) The ratio of the length to diameter of the header (or sub-header) is important too. The header length is dictated by the plot plant and from an economical viewpoint this ratio should be kept large to have a

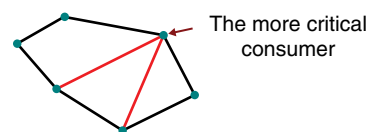


Figure 17.3 Using loop distribution to take care of a more important user.

Table 17.3 Features of each distribution system.

	Tree Distribution	Loop Distribution
Number of users	Fewer	More
Importance of users	Less critical users	More critical users
Type of usage	Intermittent users	Continuous users
"Perishable" utility fluid	Wide range of residence time of fluid in the network	Fairly constant residence time of fluid in the network
Design	Less difficult	More difficult
Chance of total fluid stoppage because of network rupture	More chance	Less chance
Applications	Default arrangement	Fire water, potable water

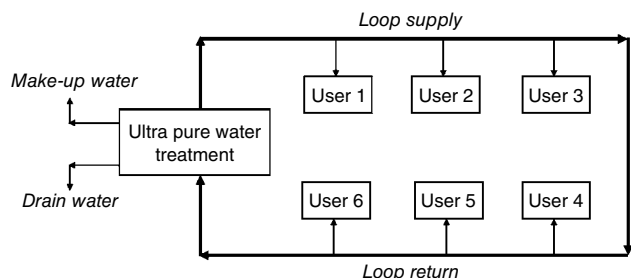


Figure 17.4 Using loop distribution for ultra-pure water.

small diameter header. However, in a low pressure utility fluid networks keeping a small ratio (large diameter) is important. While in high pressure utility fluid networks a large diameter header is not necessary. To summarize, the lower the fluid pressure, the larger the header diameter.

- 6) The old rule of thumb of “oversize the header by a factor of 2” is not only because during the design phase the utility users “expand.” It is also because a larger header diameter helps to hinder design inaccuracies caused by dictated parameters of the plot plan.

It is very common to put tees and blinds in different points of a utility network. This practice helps the plant managers to add new utility users to the utility network with minimum downtime and a limited impacted area in future.

A utility network can be arbitrarily divided into headers, sub-headers and branches. It is a good idea to put manual isolation valves (e.g. gate valves) at the beginning of sub-headers and/or branches. This help to keep a utility network functional even when there is a problem in a portion of the branch or sub-header. Such manual isolation valves in utility networks are known as “root valves” and are used to isolate a portion of a utility network from the rest of the network for repair and or inspection.

17.2.4 Placing Priority on Utility Users

In this stage we need to identify the very important utility users (if they exist) and arrange their network accordingly.

The “critical user” can be defined as a user that should never be left starving for the utility.

There are very common cases where all utility users have the same level of criticality. In such cases there is no need for specific arrangements. However, there are some other cases where a user or a group of users have priority over other users of the same utility for consumption. In such cases a specific arrangement should be implemented for this preference (Figure 17.5).

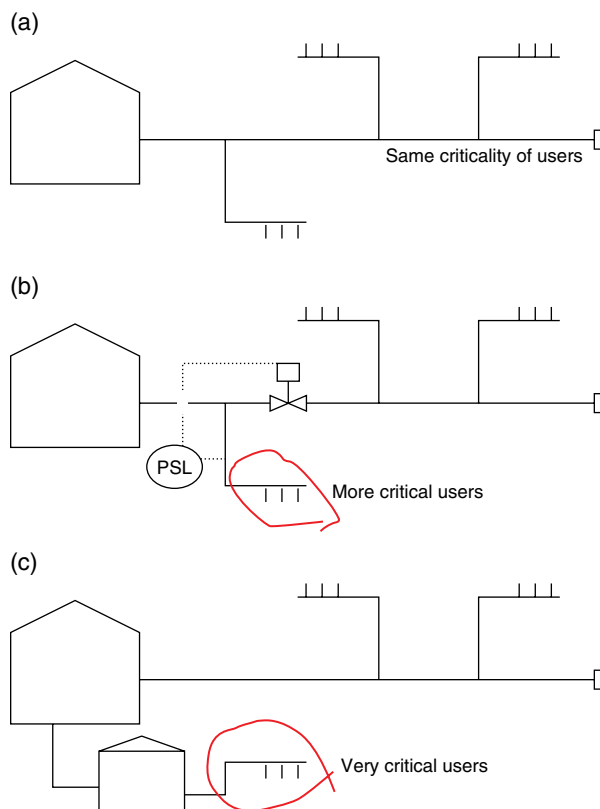


Figure 17.5 Dealing with high important users in pipe network design.

Figure 17.5(a) shows a simple utility distribution network where all the users have the same level of criticality.

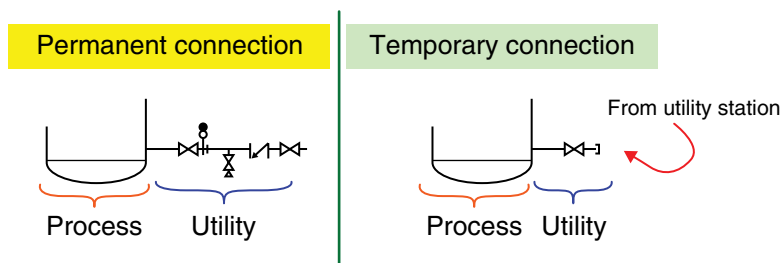
Figure 17.5(b) shows a case that a group of users is more important than others. In this case we can implement a control system (or interlock system) to disconnect the access on non-critical users from the utility when there is not enough utility for all users.

This arrangement is very common in a plant where they have one generation system for both instrument air and plant air. As instrument air is more important than plant air, this control arrangement can guarantee the availability of instrument air by scarifying plant air users.

Figure 17.5(c) shows another design for more critical users. In this design the utility comes from a separate “utility surge container” (discussed in Chapter 5). For example when an area is a big air user or is a critical air user, a local plant air receiver may be used. These local plant airs are in addition to the main air receiver inside of an air generation package.

The criticality level of users is not always carved in stone. For example it is common that we put down the criticality of a utility user to lower levels when it starts to consume a utility stream wildly. As mentioned in Chapter 14, an override control can take care of such situations.

Figure 17.6 Permanent versus temporary utility users.



17.2.5 Connection Details of Utility to Process

Utilities are used for the purpose of process. Utility streams could be connected and hard-piped to the process or could be separated from the process. A utility stream that is not hard-piped to the process ends in the “utility station” (US). Utility stations will be discussed in Chapter 18.

If a process needs to be supplied continuously with a utility stream, it should be hard-piped.

However, for a non-continuous requirement of a utility stream the utility pipe could be hard-piped to the process or only ending in the US, depending on the frequency of usage.

When a utility pipe is connected to a process pipe or equipment, adequate provisions should be considered to make sure no backflow of the process stream occurs and no utility contamination is probable. When the utility stream comes from the US, there is already a check valve installed on the stream and no additional check valve is needed near the process item (Figure 17.6).

Connecting the distribution network to the collection network will be discussed in section 17.15.

17.3 Different Utilities in Plants

There is no standard list of utilities for all plants; however, we can make a list of common utilities in plants. They include:

- 1) Instrument air (IA)
- 2) Utility air (UA)
- 3) Utility water (UW)
- 4) Potable water
- 5) Heat transfer media
- 6) Condensate collection network
- 7) Fuels
- 8) Inert gas
- 9) Vapor collection network
- 10) Emergency vapor/gas release collection network
- 11) Fire water
- 12) Surface drainage collection network
- 13) Electricity

Electricity is not a process utility. The generation and distribution of electricity is not shown on P&IDs, therefore it is not discussed here.

In the next sections, we will explain each of these utilities briefly.

17.4 Air as a Utility in Process Plants

There are at least two types of air used in process plants; they are instrument air and plant air. In some companies these two air systems are completely separate systems. This means there is one instrument air generation system and another one as a plant air generation system, and each of them has their associated distribution system. However, in some other plants they are both integrated into one system and one system provides both instrument air and plant air. In such cases, however, it should be made sure that preference is given to instrument air rather than plant air. This means if overuse of plant air starts to cause decreased pressure in the instrument air header there should be a control system to cut off the plant air branch and prevent plant air users from using plant air to make sure that instrument air is always available. This is because instrument air is more important than plant air. Instrument air is a motive gas for control valves, switching valves, and some flow meters. While plant air is the air that is used for purposes used in the plant other than those for instrument air. Plant air can be used as a motive gas for operation of an air operated pump or it could be used for operation of air cushions in silos.

The main purpose of plant air and instrument air is providing a flow of air that is dust free and water droplet free, and within good temperature.

17.4.1 Instrument Air (IA)

Instrument air is almost always necessary in process plants. IA is used to actuate control valves and switching valves remotely. Therefore IA works as the “nervous system” of a plant.

Basically, instrument air is needed wherever we have a controlled system in a plant. This is because the majority of control valves and switching valves in current industry are pneumatic. There are some non-pneumatic control valves and switching valves available, but they are currently not popular.

Generally, IA is considered one of the minimum required utilities in each plant. However, there are some cases where IA is not available in an area. For example, in the oil industry, in the well pad area (oil well location), IA is not always available because usually they are remote areas and providing IA is expensive. In such cases, some companies use the available natural gas to act as IA and actuate control valves and switching valves. The other way to avoid use of IA is by using non-pneumatic valves or motor-driven or hydraulic valves.

To be able to provide IA in a plant, an IA package must be installed in an area. The IA package converts the ambient air to IA and also provides some storage for it. Items inside the IA package depend on the required specification for the IA. Generally, IA specifications include a maximum amount of debris to make sure the debris won't plug the delicate control valve systems. The other IA specification is a low dew point; this prevents the IA from forming condensation droplets, which may also block the system.

17.4.2 Utility Air (UA) or Plant Air (PA)

Utility air, or plant air, is the air that we sometimes need in a plant for any purpose other than instruments. For example, UA could be needed for an air pad inside lime silos to facilitate discharging lime from the silo. Or, UA could be needed in a different location of the plant for short term use by maintenance personnel for their pneumatic tools.

Generally, the required specifications for UA are less strict than for IA.

17.5 Water as a Utility in Process Plants

Water, after air, is the most favorite utility in process plant. It is because of its availability and also its unique properties. The drawbacks of water as a utility is its scaling and fouling tendency.

In the old days, plants could use water without any limitations. These days regulating bodies monitor the usage of water by process plants. The process plants generally need to get a license to be able to use water because "it is a resource that belongs to all tax payers."

Water can be used as part of process as "process water," or as utility water or as potable water for personnel (Figure 17.7).

Process water is the water needed for the process. This could be as a main stream or as make-up stream.

Application of water as cooling water and water for heating purposes (e.g. to generate steam) is discussed later in this chapter.

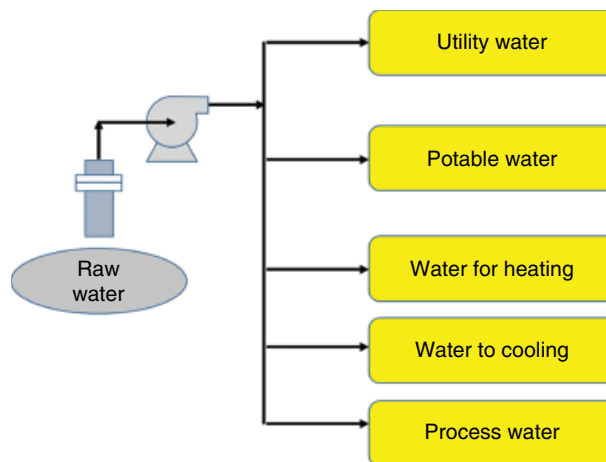


Figure 17.7 Different users of water in a plant.

Water as process water is not considered as a utility and is beyond the scope of this chapter.

17.5.1 Utility Water (UW) or Plant Water (PW)

Utility water is provided in plants for cases where water is needed for purposes other than the main process. For example, some flush pump systems need external water for cooling and lubrication purposes, and UW should be provided for such systems. The other application for UW is during maintenance, when maintenance personnel need water for washing and cleaning process equipment.

17.5.2 Potable Water

Potable water should be provided and distributed in a plant because a plant has operators and they need water. The purpose of this potable water is not only for drinking, but it could also be for safety showers and eye washing stations that are installed in different locations around the plant; the quality of the water for such systems must be similar to potable water. Potable water specifications are very strict and are usually defined by the regulating bodies in the area where the plant is located.

17.6 Heat Transfer Media

Heat transfer media includes heating media and cooling media. It is a rare case that all plant operations occur at ambient temperature. For example, a suitable temperature for a reaction could be much higher than ambient temperature, for example 150 °C. In such cases, the stream that goes to the reactor needs to be heated up. A heating medium is required. On the other hand, some

unit operations or processes require a very low operating temperature (for example 10 °C). This is an example of a requirement for cooling media. Therefore, the generation and distribution of hot and cold media in plants is very common.

Depending on different technical and economic factors, each plant may decide to have a specific set of cooling media and heating media.

Table 17.4 is a non-exhaustive list of heat transfer media and their applications.

Aerial coolers are always in competition with water coolers. Water coolers are conventional heat exchangers that are used for cooling by cooling water. Generally speaking, if the intent is decreasing the temperature of process stream down to 65–75 °C, an aerial cooler is the best option, but if the process stream is supposed to be cooled down to lower temperatures of 30–50 °C, a water cooler should be used.

Aerial coolers, however, need electricity for their fans. However, the distribution of electricity is much cheaper than the distribution of a fluid.

It is not always the case that a “fresh” heat transfer medium is used in heat exchangers. There are some cases that a “used” heat transfer utility stream is used for heat exchange purposes.

One example is cooling an oil fraction, not by cooling water, but by returned, warm cooling water. This practice could be done because not very cold temperatures are needed and at the same time using very cold stream causes some problems like setting of the oil stream. This is could be the choice for oil fractions that have a high pour point.

The other example is using hot condensate generated from steam for cooling after the steam was used for heating.

Because of the importance of steam it is discussed in depth here in Section 17.6.1.

17.6.1 Steam

Steam is one of the most expensive utilities in process plants. A steam is used for different reasons including: cleaning, purging, stripping, heat transfer, etc.

As steam is hot it can remove dirt from equipment and can be used for cleaning of equipment and instruments in dirty services during inspection and maintenance.

Steam can also be considered as a “neutral gas.” It can be used for purging of pipes and containers from non-innocent gases. A neutral gas can be used for stripping too. When steam is injected into a gas space of a non-flooded container, it decreases the partial pressure and then the dissolved gases are stripped off of the liquid surface.

It is important to know that steam is only considered as a heat transfer fluid if it is saturated. If the steam is superheated it doesn't work as a heat transfer medium.

However, transferring saturated steam over a long distance is a bit difficult. If there is saturated steam, it could be partially and then gradually converted to condensate on the route to the user. This means a portion of our valuable heat transfer medium could be lost on the way to the final user. Because of that, if the steam user, e.g. a heat exchanger, is far from steam generating system, usually the steam generating system should generate a slightly superheated steam. It is wiser to transfer a slightly superheated steam, which will hopefully be converted to saturated steam when it gets to heat exchanger. If not, then the superheated steam should be the desuperheated before directing it to the heat exchanger, as stated in Chapter 11.

Desuperheating a steam stream can be easily done by a desuperheater. A desuperheater is basically a piece of equipment that converts superheated steam to saturated steam by injecting an adequate amount of water to the stream.

An infinite number of types of steam with various pressures and temperatures can be generated in each process plant, but to make things simple usually only few different types of steam in process plants are generated. It is very common to see the three-type steam program in process plants. They are low pressure steam (LPS), medium pressure steam (MPS), and high pressure steam (HPS). Each of these three steam streams are for specific applications.

Table 17.5 gives the features of these three steams.

Distributing steam through the piping network needs some consideration. The main issue is heat loss and generation of condensate in steam pipes. As steam is in gas phase and condensate is in liquid phase, the best efforts should be considered to handle these two phases of fluids carefully in pipes. The potential result of flowing steam and condensate together is “steam hammering,” which may damage the

Table 17.4 Heat transfer media and their applications.



 		
Users	<ul style="list-style-type: none"> • Process heating (heat source) • Building heating • Anti-freeze protection of process lines (tracing) 	<ul style="list-style-type: none"> • Process cooling (heat sink) • Pump seal flush plans
Type of heat transfer media	Heating media: Steam, hot glycol, non-water-based heat transfer liquids	Cooling media: ambient air, cooling water, cold glycol, refrigerants, cryogenics

Table 17.5 Some typical utility users in process plants.

Steam type	Pressure	Temperature	Applications
LPS (low pressure steam)	100–200 KPag	120–135 °C	In utility stations, In heat exchangers (steam heaters)
MPS (medium pressure steam)	1000–2700 KPag	185–230 °C	In heat exchangers (steam heaters)
HPS (high pressure steam)	>4000 KPag	>250 °C	In steam turbines, as process steam

piping system. Steam hammering is a phenomenon that may happen in steam–condensate two-phase flows where the steam carries the pockets of condensate and smashes them into pipe obstacles like elbows, partially open valves, etc. and damage them.

This concern can be solved by placing steam trap on pipes with a chance of receiving steam–condensate mixed fluids. This steam trap prevents steam from passing. As steam traps are available only up to certain sizes (say up to 6"), if the pipe is large enough, instead of a steam trap, a small steam–condensate vessel with level control could be placed.

The generated condensate along the length of a steam carrying pipe should be removed from the steam as soon as possible. Therefore steam traps are placed along the route of the steam transferring pipes at certain intervals and the steam pipe should be installed so that they slope toward the steam traps. The steam trap interval is case specific and depends on several factors including pipe size, the effectiveness of pipe insulation, and the difference between steam temperature and ambient temperature.

Steam traps are connected to the steam pipes through a device named a drip leg. Wherever we need to put drip leg, it should be from the bottom of the steam pipe. The detail of drip legs are not always shown on main P&IDs. If the detail of drip legs is not on the main P&IDs they could be on auxiliary drawings or even in piping documents.

After all these consideration there could still be some condensate in the steam pipes. To avoid directing the generated condensate to the steam branches, all the steam branches should be taken from the top of the steam header. This can be covered as a note in P&IDs.

As the bad reputation of valves in steam services is widespread, care should be taken to use suitable valves.

17.7 Condensate Collection Network

The condensate collection network is the mating system for the steam distribution network. The condensate collection network collects the generated condensate from

intentional and/or unintentional conversion of steam. This conversion happens because of heat transfer. The heat transfer could be intentional in steam heaters or unintentional in steam transfer pipes.

As there are generally three types of steams: low pressure, medium pressure, and high pressure, there could be three corresponding condensate collection systems, including low pressure condensate, medium pressure condensate, and high pressure condensate.

17.8 Fuel as Utility

Fuels are needed wherever there are burners in the plant. Burners can be used in different pieces of equipment including furnaces, boilers, some HVAC equipment, and others. Burners are a piece of equipment that has the duty of generating a flame and also maintaining it.

Fuels could be gas, liquid, or solid form.

A famous solid fuel is coal. If the usage of coal is justified it is generally used in a way such that distribution of it is not necessary. Therefore, there is no such thing as coal distribution system in an extended scope.

Here we cover two main types of fuel: fuel gas and fuel oil.

Generally speaking fuel gas is the preferred fuel for burners rather than fuel oil or coal.

17.8.1 Fuel Oil

Fuel oil is another option as fuel in burners. Fuel oil is less favorable than fuel gas for burning.

17.8.2 Fuel Gas

Gas burners need to be supplied by streams of fuel gas and air. In some plants, there is one network to distribute natural gas both for blanketing purposes and as fuel gas (if natural gas is used as fuel gas). However, in some other plants there are two separate networks: one for blanketing gas distribution and the other for fuel gas distribution. This could be because the blanketing gas specifications required by the client could be different from the fuel gas specifications. For example, in some

plants the fuel gas specification requires the addition of an odorizing compound to the natural gas to make the detection of leakage easy for operators.

17.9 Inert Gas

“Inertness” could be a confusing adjective here. “Inert” here means a gas that would not be involved in any reaction with the liquid contents, nor with the equipment body material. As a fundamental requirement, an inert gas should not be toxic to humans and should not be flammable.

The best and the most expensive inert gas is nitrogen gas, but this gas is usually not economically viable for use in processing plants. One common application of nitrogen gas as an inert gas is in the food industry. The other expensive gas used as an inert gas is carbon dioxide.

Another inert gas that is very common in the oil industry is natural gas. The main composition of natural gas is methane, which is a very stable chemical and can be considered as an inert gas. However, one concern that may be raised is the flammability of methane or natural gas. However, this concern is not valid. There is a chance of combustion only when all elements of the fire triangle are available (air, flammable material, and heat). When the oxygen content is minimal, the mixture is too rich to inflame.

17.9.1 Blanket Gas

In some atmospheric tanks, there is a need to provide a specific atmosphere at the top, empty space of the tank above the liquid level. This can be done by blanketing (or padding).

Leaving that top space without any specific atmosphere would provide an opportunity to have a space full of air (oxygen) and vapors from the liquid. This could be dangerous for the liquid or for the tank. In the majority of cases, the existence of oxygen in that space would promote corrosion inside the tank, which is not a good thing. In some other cases, leaving that space with oxygen and flammable vapors from the liquid would create a flammable mix, ready to be ignited and cause the tank to explode.

In still other cases, the conventional air atmosphere can degrade the liquid. For example, some petrochemical plants have tanks full of liquid monomer, which is stored to be sent to the polymerization unit; however, the existence of oxygen in the top space of the tank would trigger the polymerization reaction of the monomer in the tank, rather than in the polymerization unit. A polymerization reaction in a tank is not good because it is not designed to handle this. In this case, again we need to create a specific atmosphere instead of a conventional air atmosphere in the top space of the tank. Providing a specific atmosphere inside the top space of the tank can be done by introducing a specific gas into that space.

In the hydrocarbon industry, natural gas is the most common blanketing gas. The only concern regarding the application of natural gas as a blanketing gas is some solubility of methane into the liquid. If dissolved methane in the stored liquid is acceptable, then natural gas can be used.

If in a plant there is only one tank requiring blanketing, then there is no need for a “blanketing gas distribution network.” However, if there is more than one tank that needs blanketing gas, a blanketing gas distribution network should be installed in the plant, and this can be considered as another utility network.

17.9.2 Purging Gas

Inert gas is used where it is needed to push out a gas or vapor from the piping or equipment. One famous purging gas is nitrogen.

A gas or vapor would be the target for removing gas from a space for different reasons. Air or a hydrocarbon gas may need to be pushed out of a system to prevent creation of a flammable mixture. This operation is named purging. The operation may be needed after the plant shutdown or before the plant start-up.

Providing inert gas for burners and fired heaters is very common.

While steam cannot be used for blanketing it is common to use it for purging purposes.

17.10 Vapor Collection Network

A vapor collection network is designed to collect vapors out of process units that are not categorized as “emergency” vapors. The most famous non-emergency vapor producers are tanks. Tanks generate vapors because of their normal breathing.

A vapor collection network is a set of pipe routes to direct the vapors to a vapor recovery unit (VRU) (Figure 17.8).

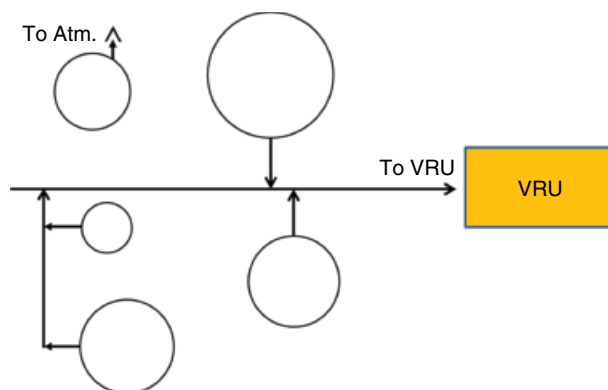


Figure 17.8 Vapor collection network.

The non-emergency vapors may be released to atmosphere if they are not harmful and/or the recovery is not economical. Such vapors are more commonly released to atmosphere directly from each single vapor generator or through a “cold vent” at the end of the vapor collection network.

The emergency vapors should be released to an “emergency vapor/gas release collection network.”

17.11 Emergency Vapor/Gas Release Collection Network

The duty of an emergency gas/vapor release collection network is collecting gas and vapor release from pressure relief devices or blowdown systems to an emergency release handling system. The most common type of emergency release handling system is flare. This is the reason that this system is generally called a “flare collection network” (Figure 17.9).

A flare collection network has different specific features. First of all flare networks cannot handle large amount of liquids in them. Then they may need intermediate (local) KO drums if the liquid content of released fluids from safety devices is high. However, there is always at least one main KO drum available just before the flare. To remove the liquid content (e.g. condensates) effectively then branches, sub-headers and header should be sloped toward the local KO drums and main KO drum.

When the set pressure of pressure relief devices is high the “kicking” to the piping network could be high and non-acceptable. One solution is using long radius elbows rather than standard elbows and using 45° or 60° tees instead of a 90° tee.

The manual valves should be eliminated or at least “locked in the open position.” Such provision decreases the chance that the pipe is closed by a careless person.

The automatic valves on the flare collection network are not a good provision, even with “fail open” on them, because they may “jam” in the closed position.

Valves in a flare header should be of gate type with stems in the vertical downward position to prevent them collecting debris and jamming.

All these points could be covered under notes in P&IDs.

17.12 Fire Water

Even though there are many different chemicals that can be used to fight against fire and suppress fire, arguably water is the most common type of chemical that can be used for the purpose of fire suppression. Therefore a fire water system is almost always an integral part of process plants.

Fire water systems are much regulated systems and the design and construction aspects of them are generally proposed by jurisdictional codes. To design an acceptable fire water system an engineer should consult with the local fire codes. We, however, here just cover the general technical notes regarding P&IDs of fire water systems.

The block flow diagram (BFD) of a fire water system can be seen as four different blocks/functions: fire water storing, water pressurizing and mobilizing, water distribution, and water discharge. A BFD can be seen in Figure 17.10.

Fire water storage could be an above ground tank to store water for few hours to fight against fire in a process plant. Sometimes, this fire water storage tank is not a dedicated tank and it could be combined with other tanks. The reason for this practice could be the less frequent usage of fire water storage tank that makes it difficult for investors to assign specific capital for a tank that may be used once in every several years. One common option is having a raw water/fire water storage tank. In such a provision, a tank is placed and used for the two purposes

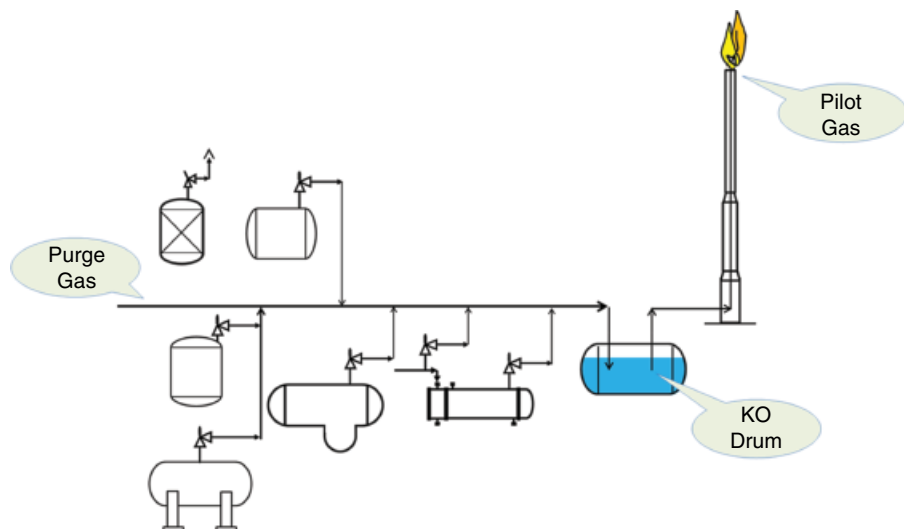


Figure 17.9 Emergency vapor/gas release collection network.

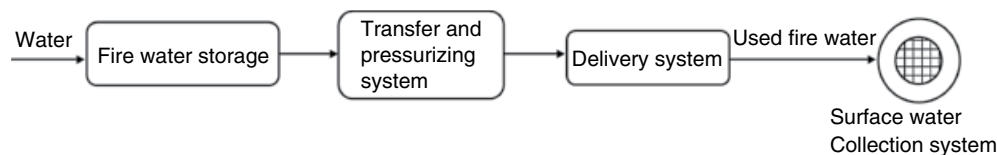


Figure 17.10 BFD of a fire water system.

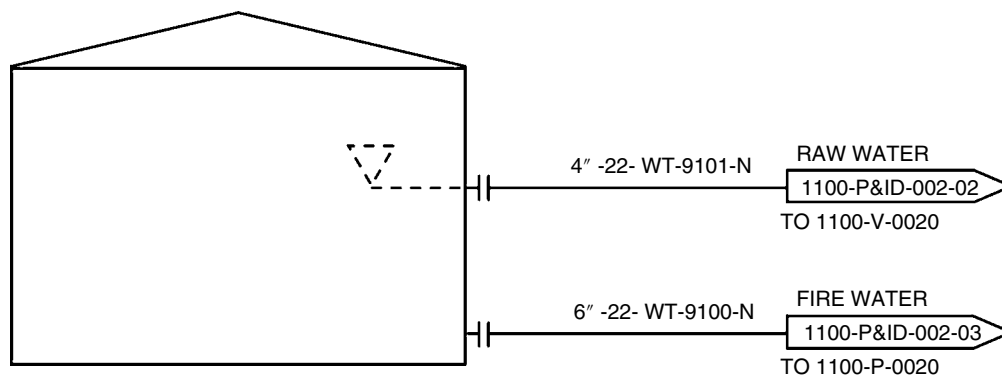


Figure 17.11 Arrangement of shared fire water and raw water tank.

of storing raw water and storing fire water in a process plant. However it should be make sure that in no situations should raw water users consume the water from the shared tank in a way that there is no available fire water in the tank. To provide such peace of mind a specific arrangement should be used. Figure 17.11 shows one of the available arrangements inside a shared raw water/fire water storage tank to make sure there is always available water for firefighting purposes.

A water pressurizing and mobilizing system is generally done through a set of pumps.

To provide high reliability for the system and to make sure that there is always at least one pump available, pumps are generally operated in spare/operating mode. It means there is always spare capacity available for pressurizing fire water in the network. The other provision to increase the reliability of the system is using different pump drives. For example one pump can be driven by an electric motor and the other pump by a gas engine or diesel engine.

There is one other issue related to pressurizing fire water. In large water distribution systems like a fire distribution system, the pressure of the fire water in far locations of the network could be low, and lower than the adequate pressure for effective water discharge. The reason for this is that pumps are stopped and not working when there is no fire water released into the plant, therefore, because of different reasons including leakage, there could be some low point in the fire water network. Such low pressure points in a fire water system cannot be afforded because a reliable fire water system should be available every time when there is a fire in the plant. Therefore we need to make sure the fire water system is always pressurized with the pressure that is required to discharge the water on the fire.

To support such idea, we generally put another pump in the fire water system that we name a jockey pump. Jockey pumps are generally centrifugal type pumps that are in the fire water network and near the main fire water pumps but much smaller than them. The capacity of the jockey pump could be even one to ten percent of the main fire water pump. This jockey pump works best in a control loop that whenever the pressure of the system decreases beyond a specific number it can turn on and start to pressurize the system again (Figure 17.12).

A fire water distribution system is a network that provides fire water to different locations of a process plant. To make sure that fire water is always available at every point with a predetermined pressure the main header of a fire water network is in the form of a ring. The fire water ring is the main header of the water network that is around the process plant.

The fire water network could be underground or an above-ground piping network.

A fire water network should have the minimum number of valves on it to prevent unintentional closing of the valves by a careless person. Valves on the fire water network are mainly on/off valves and not throttling valves. If there is an on/off valve on a fire water network it should be locked open (LO) or car seal open (CSO). The valves should be of “non-rising stem” (NRS) type to make sure that when the valve is open it is easily visible and recognizable by a person in the vicinity of the valve. When the fire water network is underground, it is a good practice to use “post indicator valves,” which are special long bonnet valves. Using valves-in-a-pit should be avoided to facilitate frequent checking and testing of the valves on fire water systems.

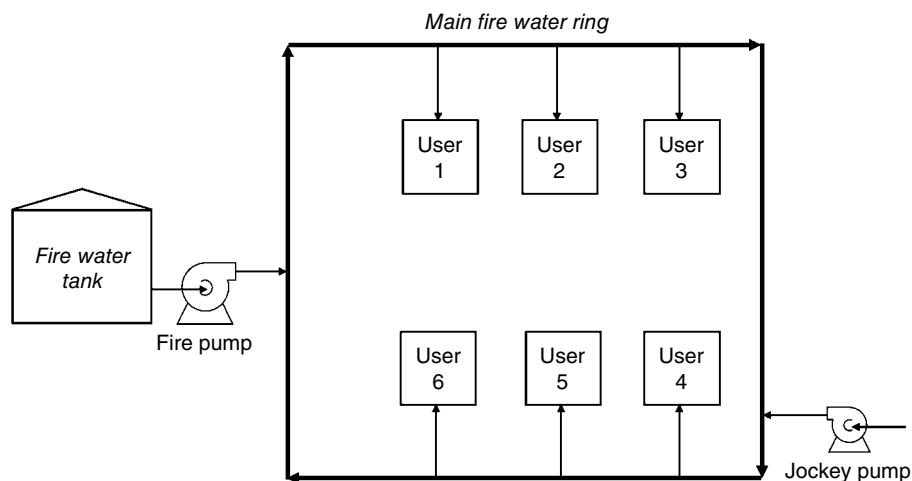


Figure 17.12 Fire water loop.

If there is a need for any type of control valve we generally put a mechanical control valve, or a more common term, regulator. The reason that we don't like to put conventional control valves with a control loop is because of their dependency on the instrument air in the plant and we are not sure if the instrument air in a plant is available during a large fire in a plant. We may have pressure regulators or back pressure regulators or flow regulators in a fire water network, but we need to make sure they are failed open types (FO). For pumps in a fire water system there is a device for minimum flow pipes that is controlled, not by a conventional control loop, but through an automatic minimum flow valve.

If there is a chance of freezing, the system should be heat traced to make sure no chunk of frozen water plugs the network.

Because a fire water system should be tested very frequently for its availability the system should be provided with adequate test points. The quality of water doesn't need to be very high but at least it should be with a quality that supports minimum plugging, clogging and corrosion even in static condition.

Water discharge devices in a fire water system could be sprinklers, hose over the hose reel, water gun, hydrants, etc. and after releasing fire water there should be system to collect the water to prevent wasting it.

17.13 Surface Drainage Collection Network or Sewer System

There are three different waste water streams in a process plant:

- 1) Industrial waste water (process wastewater)
- 2) Municipal waste water generated by operation personnel
- 3) Surface drainage.

Each of these three waste waters should be handled through different concepts but not necessarily through different waste water treatment networks/plants.

The treatment of process wastewater is beyond the scope of this book because treatment of waste water from a process plant is different for each type of process plant. The treatment of waste water from an oil refinery could be different than treatment of waste water from food processing plant and could be different than electroplating workshop.

Treatment municipal waste water is again beyond the scope of this book.

The last type of waste water is discussed partially here.

There are at least four main types of surface waters in plants: storm water, floors washing water, draining dirty liquids, and used fire water. All of these streams are classified as waste water and should be collected in sumps.

Sumps are containers that receive different surface water from the plant. Because these water streams don't have enough pressure generally speaking, the sumps should be underground to be able to receive gravity flow surface waters easily.

Storm water is basically the water run-off from rain and other precipitations.

Floor wash water is the water that is released on the surface by operators when they wash the floors.

Dirty liquids come from draining and cleaning of equipment or when process equipment is emptied to start doing inspection or maintenance.

Used fire water is a very large contributor in surface drainage. It is created during a fire or shortly after that.

Not all these streams exist in every area of the plant.

For example storm water is not present in indoor industrial areas. For outdoor industrial areas that storm water could be present, it could be in two forms. In each process plant, outdoor areas could be in two main forms: form 1

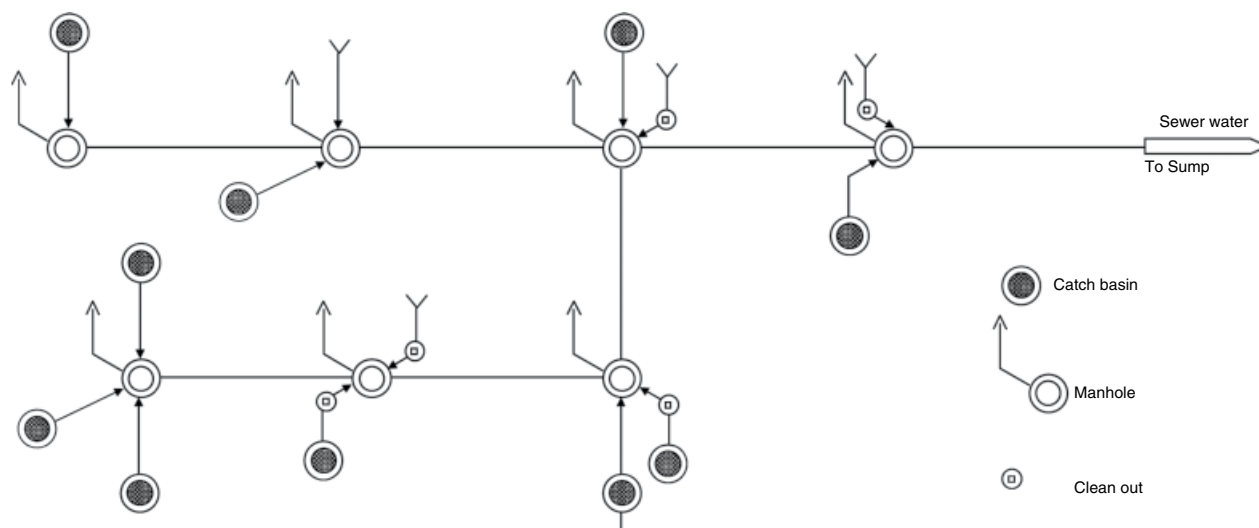


Figure 17.13 Surface drainage collection network.

is the ground where there are concrete pads and equipment installed on them. The form 2 is soil ground. It could be assumed that rainwater that comes on pad areas is contaminated storm water because process equipment is installed on those paths and there is a chance of chemical spillage. However the rain on the soil area could be considered as non-contaminated storm water.

The “footprint” of a surface drainage system is very small on the main P&ID. It could be limited to notes near discharge points as “to drain system” with or without a symbol representing it. The end of the system, or “sumps” could be shown on the main P&ID or auxiliary P&IDs. The collection network can be handled in different ways. In some plants they don’t show it at all on P&IDs with the logic that “they are civil engineering considerations and we don’t show them.” In some other plants they show them on auxiliary P&IDs.

Such a surface drainage collection network is shown on a P&ID as shown in Figure 17.13.

Surface wastewater is generally collected through a network of tranches. However, there are cases that it is done through pipes too. The pipes could be laid down in trench or buried as underground pipes.

The minimum size of pipes for surface wastewater collection could be considered as 2”.

Figure 17.14 shows a surface drainage collection network with pipes.

In some cases dumping the liquid on the open-top trenches involves hazard risks. For example if the liquid is very flammable or toxic. In such cases the drain should be hard-piped toward a closed sump system.

The surface drainage is generated by two types of sources: a point source and a surface source. An example of a point source is the draining nozzle of a vessel. An example for a surface source is washing water. It is

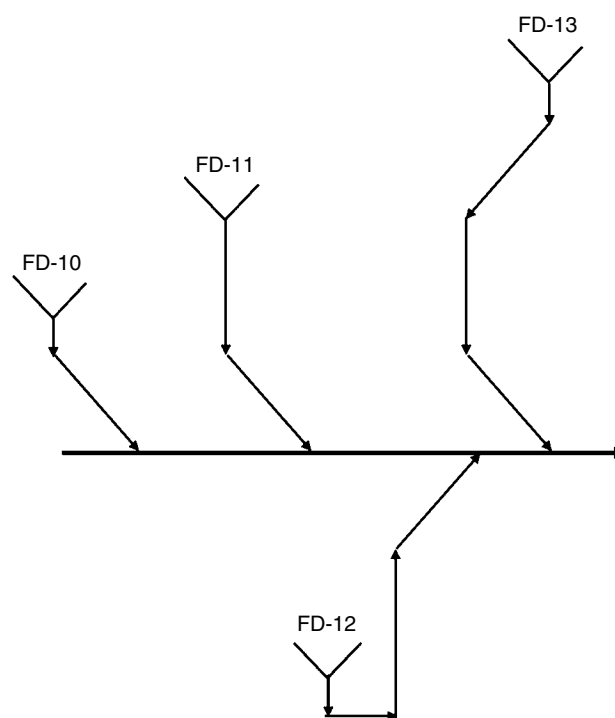


Figure 17.14 Piping network as a surface drainage collection system.

fairly easy to direct the water from a point source to the trench, it just needs to pipe the point source to the trench. For surface sources, however, the floor should be sloped toward the collection network. Sometimes there is a need for several sumps for a specific area just because sloping is not possible to be implemented in the area.

In some plants, there is need for a separate surface drainage collection system and separate sumps named

segregated systems. This could be because of technical or regulatory requirements.

Examples are:

- Oily storm water and non-oily storm water sump and network
- Acid and base sump and network
- Chemical sewer system and non-chemical sewer system
- Open sump and network system, and closed sump and network system.

Segregated contaminated storm water and non-contaminated storm water sump and networks are very common because of jurisdictions on process plants that do not allow them to discharge their contaminated storm water to the environment or public sewer system before treatment. A famous example in oil refineries is that oily storm water and non-oily storm water is very common and should be handled separately.

In areas that handle strong acids and strong bases, they may have a dedicated acid sewer and base system to direct the spilled or drained acid and base through a specific collection system and to specific sumps. This requirement is because of the very exothermic nature of the reaction between strong acids and strong bases. There are some cases that such two separate systems have one common sump though. However, precautions should be considered in the design of a common sump receiving acids and bases.

When the surface drainage liquid contains low flash point liquids the collecting system and the associated sump should be “closed” to prevent vaporization and the generated vapors catching fire. Then a closed sump system is needed that comprises a pipe network (rather than a trench network) and an underground pressure vessel (rather than a non-pressure vessel or an open top container) as the sump.

17.14 Utility Circuits

In this section we only try to give an bird’s eye view of utility circuits.

There are two main groups of utilities: supplying utilities and disposing utilities. The supplying utilities provide energy and non-process material to equipment in process plants. The disposing utilities move away the generated “waste” from equipment.

The famous energy utility is electricity.

Supplying utilities are something that can be easily available freely or with an affordable price. The primary process utilities are air (in the gas-phase group), water (in the liquid-phase group), and coal (in the solid-phase group). The effort is if a piece of equip-

ment needs a utility, one of these three substances – air, water, and coal – is used as much as possible. If for whatever reason these utilities cannot satisfy our requirement we may need to deviate from them and use other utilities.

The heart of supplying utility circuits are generation, treatment, or preparation units. All the utilities after the generation should be equipped with a sort of storage container. The reason is we don’t want to design a utility generation system based on the maximum utility usage in the plant. We generally design the capacity of utility generating system based on average consumption and then by providing by placing a storage container at the end of utility generating system and at the beginning of utility distribution network we address the short term higher consumption rights.

The heart of disposing utility circuits is ultimate disposal units.

If possible, it is very attractive that two compatible utilities, one on the supplying side and the other one on the disposing side, are combined together to minimize the treatment system and ultimate disposal systems. Such combined systems can be named “mating systems” otherwise the system could be a once-through system.

Mating systems are fairly closed recirculation systems.

An example of a mating system is a steam–condensate circuit and an example of a once-through system is an instrument air system.

Mating system utilities are more attractive because they are more economical than once-through utility systems. If a type of treatment is needed, in once-through systems the full flow needs to be treated while in mating systems only the make-up stream needs to be treated, which is a less expensive option.

These concepts are shown in Figure 17.15.

Another example of a mating system is a glycol system in which a heating glycol circuit is connected to a cooling glycol circuit.

17.14.1 Air Circuit

The air circuit starts with sucking air from ambient air and finishes with wasting the “used” instrument air or plant air to environment (Figure 17.16).

The block flow diagram (BFD) of an instrument/plant air could be as shown in Figure 17.17.

The raw material for an instrument/plant air system is definitely ambient air because ambient air is a free resource around us. However, the ambient air has low pressure, it may have dust, and it carries some humidity. Therefore we need to install a compressor, filter and dehydrator to remove all the harmful impurities in the ambient air to be able to convert it to instrument/plant air.

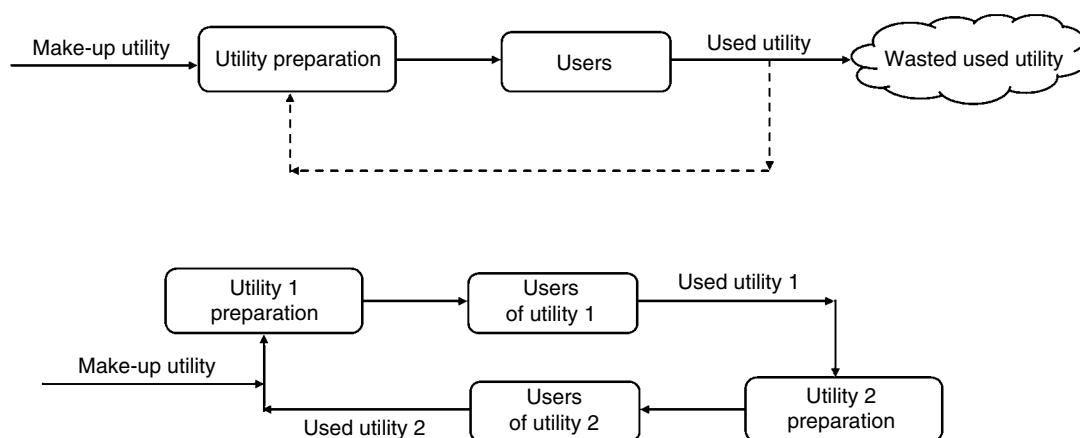


Figure 17.15 BFD of once-through and mating utility systems.

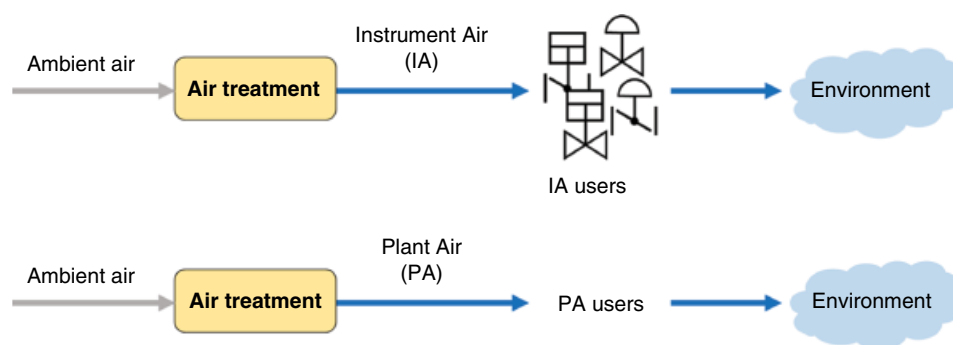


Figure 17.16 Instrument air and plant air route.

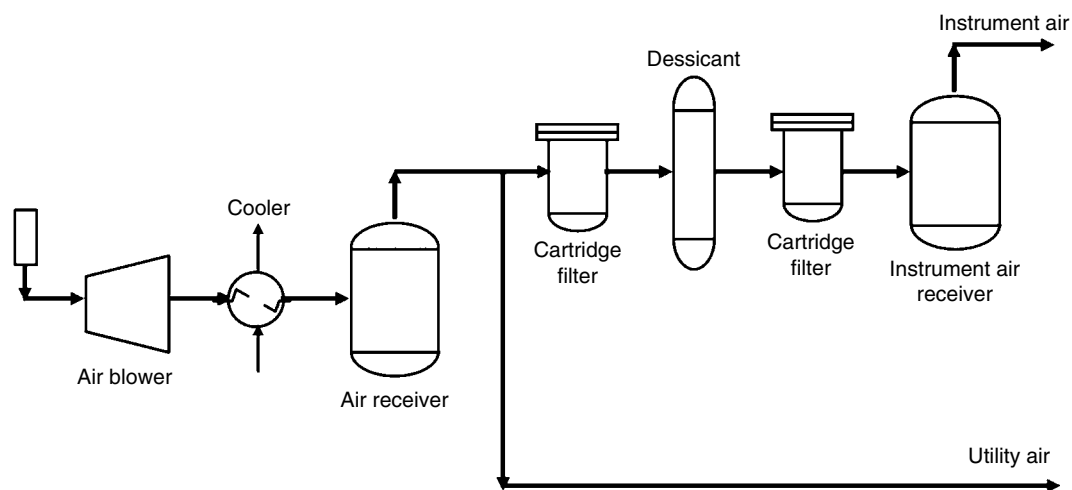


Figure 17.17 Air circuit BFD.

The type of compressor depends on the consumption and could be a centrifugal type air compressor or of a positive displacement type. Generally, the required pressure for instrument air users is a pressure between 6 to 15 psig. We usually provide instrument air for instrument air users at a

pressure around 45 psig. Because of that, the instrument/plant air system should provide a pressure more than that to be able to overcome the resistance in a route between the instrument/plant air generator and the final instrument air user. This pressure could be a pressure around 50 psi.

To remove a small amount of dust or suspended solids from air, a vessel of cartridge filters is enough as the dust concentration in water in air is low.

To remove humidity and water droplets from instrument air, a desiccant tower can be used. Desiccant towers contain water absorber beads.

Air receivers are the vessels to work as a surge vessel for an instrument air system. It has been seen that sometimes we have one air receiver at the end of instrument air system, the second air receiver at the middle of instrument air receiver and several more instrument air receivers in some areas that there are large instrument air users as “local air receivers.”

17.14.2 Steam–Condensate Circuit

In plants, steam and condensate circuits are connected together as a mating systems (Figure 17.18).

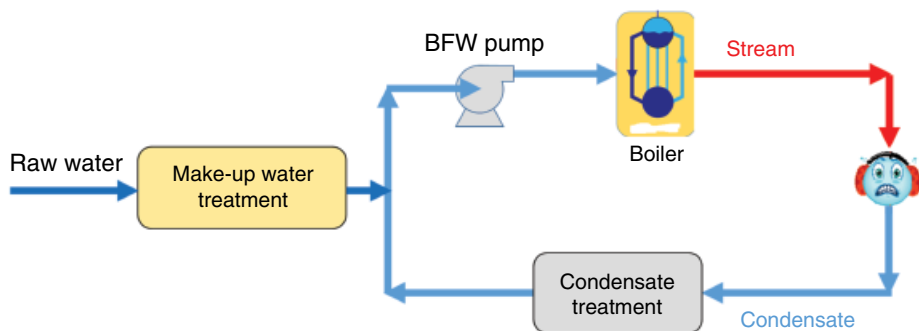


Figure 17.18 Steam–condensate circuit pair.

Only the make-up water needs to be treated. The type of treatment depends on the pressure of the boiler or the required pressure of steam out of the boiler. The treatment could be as simple as hardness removal and de-aeration, up to complicated demineralization.

The boiler feed water needs to be pre-heated before it gets directed to the boiler to minimize the thermal stress on boiler components.

Condensate may also need treatment depending on the type of equipment that generates condensate. For example in power plants “condensate polishing” is a very common unit to remove small amounts of oil contaminant in the condensate.

Steam generation is done in boilers or steam generators. Drum type boilers are the most common steam generators. Drum type boilers can only generate saturated steam (Figure 17.19)

Once-through steam generators (OTSGs) can generate saturated or superheated steam.

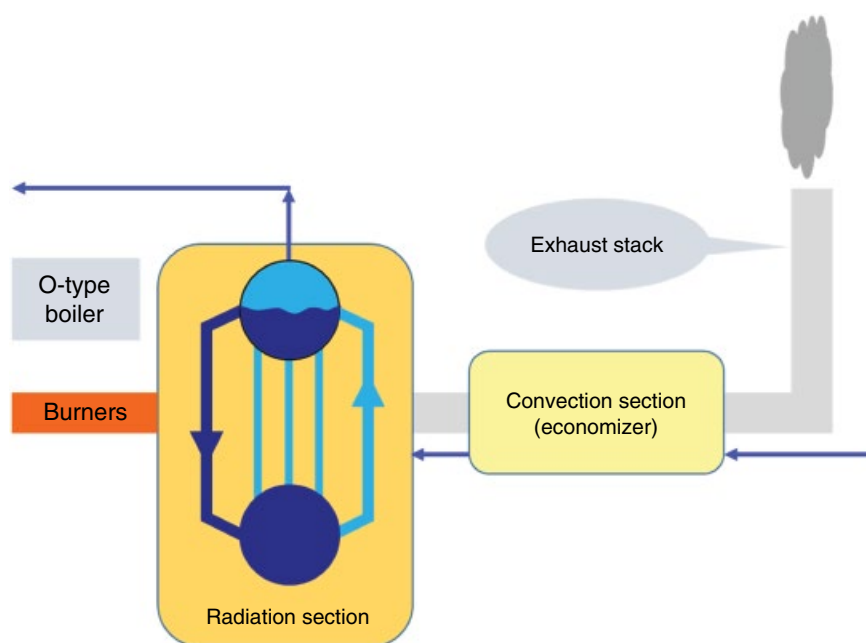
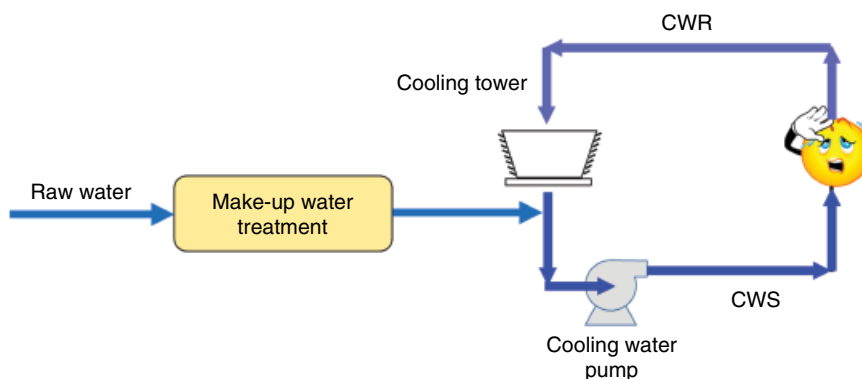


Figure 17.19 Steam generation.

Figure 17.20 Cooling water circuit.



In utility plants, generally the highest required steam is generated and then this stream can be converted to all other steams with lower pressure and temperature by injecting a suitable amount of water in the high temperature/pressure steam.

17.14.3 Cooling Water Circuit

Cooling water is provided for plants for cooling of process streams and units. The source of cooling water could be ground water, surface water, sea water or even treated waste water.

The treatment of cooling water depends on the water analysis and is beyond the scope of this book. However, it could be said that the treatment comprises injection of different chemicals into recirculated cooling water. Therefore P&IDs of a cooling water preparation system are generally several chemical injection systems.

A schematic of a cooling water circuit is shown in Figure 17.20.

A glycol heat transfer utility system could be considered as an “upgraded” version of steam or cooling water systems. In some process plants a glycol heat transfer utility is needed rather than a steam system or even a cooling water system. In plants, heating glycol media and cooling glycol media circuits are mating systems (Figure 17.21).

As the circuit is closed, expansion drums are needed to handle the expansion of liquid after increasing the temperature.

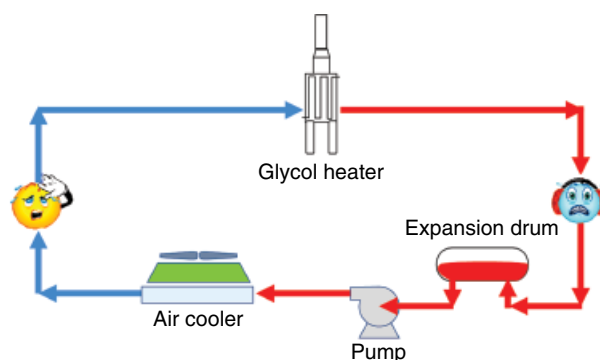


Figure 17.21 Glycol circuit pair.

otherwise natural gas can be brought from a third party, mainly through a pipeline.

If natural gas is produced in a plant with different qualities it is better to mix them together before using them as a utility to ensure the constant quality of the natural gas toward the utility system. In these cases there could be a “mixed gas drum” to mix all different natural gas streams within the plant.

The usage of natural gas can be split into two main applications: for burning and other applications.

The natural gas for burning is named “fuel gas” and the natural gas for other applications is named “utility gas.”

Fuel gas is uniquely used in gas burners.

Utility gas can be used for blanketing, purging (e.g. flare header) and other applications.

17.14.4 Natural Gas Preparation System

Natural gas can be supplied from the natural gas pipeline.

In hydrocarbon industries natural gas may also be generated within the plant and it can be supplied from the plant itself.

If natural gas is produced within a plant as the main product or by-product, it is more economical to use it,

17.15 Connection Between Distribution and Collecting Networks

It has been shown that there are some utilities that require a collection system. In such cases, the combined distribution and collection networks could be set up as seen in Figure 17.22.

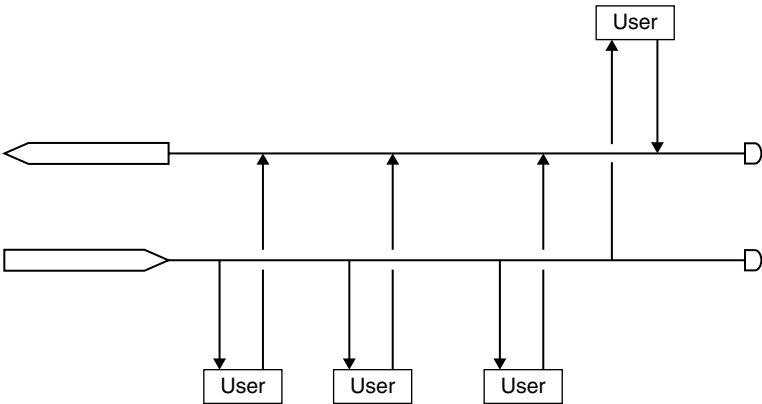


Figure 17.22 Distribution and collection network of a utility.

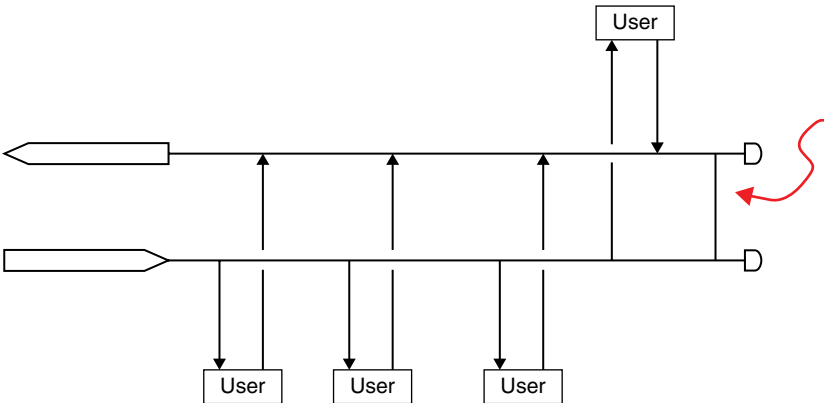


Figure 17.23 Connecting pipe between distribution and collection networks.

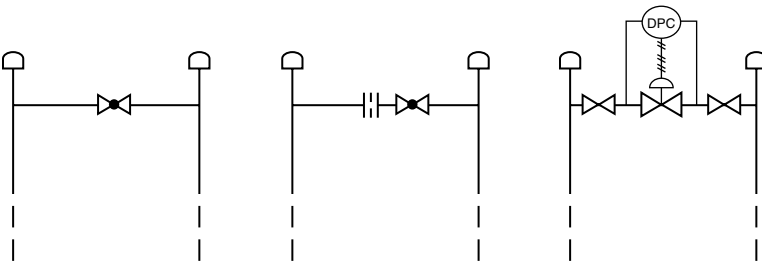


Figure 17.24 Connection between distribution and collection networks.

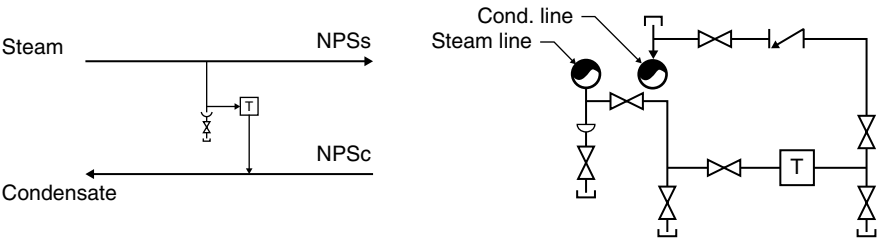


Figure 17.25 Connection between steam and condensate networks.

There is generally a connecting pipe between the distribution network and the collection network (Figure 17.23).

A controlling mechanism on the connecting pipe should be implemented to make sure that the pressure of the distribution network is always higher than that of the collection network, and that the flow is always from the distribution network to the collection network. The controlling mechanism could be anything from something as simple as a globe valve, all the way up to a full control loop based on pressure differential (Figure 17.24).

However, the connection could be more complicated in some other cases. One important example is a connection between a steam distribution network and a condensate collection network (Figure 17.25).

Wherever there is connection between steam network and condensate network, a steam-condensate separator, or a steam trap, should be installed. The size of the condensate pipe could be the same size as the steam pipe or smaller than it, and down to one third of it. In some companies the size of condensate pipes are limited to 6" to 10".

Part V

Additional Information and General Procedure

Part 5 has two chapters, Chapters 19 and 20.

In Chapter 19 several general procedures are provided. First of all a general methodology is provided for P&ID development of a new item (not familiar for the designer). Then a general procedure for P&ID checking and reviewing is provided.

At the end, the quality required for each stage of P&ID is provided.

Chapter 20 is devoted to several P&ID examples.

18

Ancillary Systems and Additional Considerations

18.1 Introduction

In this chapter we cover some systems that couldn't be categorized in the previous chapters.

18.2 Safety Issues

The main purpose of taking care of “safety” in process plants is preventing injury, and in its worst case, preventing death. Safety issues should be addressed in all aspects of process design and also P&ID development. The first step in upholding safety is understanding hazards and their relation to injuries.

Safety should be considered in all P&ID development activities. No effort is made to make this section a fully exhaustive section on safety of process plants.

18.2.1 Different Types of Hazards

Hazards can be arbitrarily classified into three groups based on the initiators: mechanical hazards, chemical hazards, and energy hazards.

Mechanical hazards are the hazards caused by mechanical systems and devices. The examples are: impacts, penetrations, compressions, rolling-overs, falling (including slipping and tripping).

Chemical hazards caused by chemicals in different forms include: liquid, gas, vapor, fume, and dust.

Energy hazards are caused by light, optical radiation, contact with hot or cold surfaces, or noise.

Hazards caused by biological matter are not generally categorized as safety hazard but are known as health hazards.

18.2.2 Hazards and Injuries

Before the injury all effort should put into preventing an injury by reducing the risk of injury. After the injury, all effort should be aimed at mitigating and limiting the consequences of an injury.

Table 18.1 shows these two concepts.

In the left column – or before the accident – all the efforts are to minimize the hazard to prevent the injury. In the right column – or after the accident – the injury has happened and efforts should focus on minimizing the extent and breadth of the injury.

Let's start with the left column.

In the scope of preventing injury what can be implemented during the design of a plant is firstly reducing or eliminating the hazard. Eliminating the hazard can be done by passively eliminating hazardous matter. The passive prevention of injury generally goes into the deep concepts of process and generally cannot be implemented in the plant during P&ID development. Such strategies can be implemented in the BFD (block flow diagram) or PFD (process flow diagram) development stages.

During the P&ID development stage of projects, active methods – or placing barriers – is the main strategy to prevent injury and reduce hazards.

The third strategy to reduce hazards is implementing rules and standard operation procedures and forcing operators to follow them. This the weakest way of dealing with hazards and also doesn't have any impact on P&IDs.

Therefore we focus only on the first item of preventing injuries actively by “masking” hazardous matters.


In the right column we only put few of the actions. Out of these actions, providing safety showers and eye washers have P&ID footprint and will be discussed here.

18.2.3 Mechanical Hazards

There are different “guards” available to protect personnel against mechanical hazards. However they are generally not shown on P&IDs. The examples are different types of “machine guards” including shaft guards, belt guards, coupling guards, etc.

The majority of mechanical hazard barriers are offered by equipment vendors.

Table 18.1 Hazard and injury.

Hazard	Injury
	
Preventing the injury	Mitigating the injuries' impacts
<p>The hazards can be controlled through three different strategies:</p> <ol style="list-style-type: none"> 1) Reducing or eliminating the hazard through engineering passively. 2) Reducing the hazard actively by putting barriers between workers and hazards. The barrier could be on the equipment side or operator side. The operator-side barrier is a type of "personal protection equipment" (PPE). 3) Reducing the hazard through establishing rules for work practices and implementing them. 	<p>The injury consequences can be mitigated through different methods including:</p> <ul style="list-style-type: none"> • Providing first aid • Providing safety showers • Providing eye washers.

18.2.4 Chemical Hazards

Chemical hazards can be reduced by placing, again, some sort of guard on the areas where there is chance of exposure. For example for aggressive material, "splash guards" could be placed on flanges of piping of aggressive liquids.

"Splash guards" are generally stated in a note on the P&ID.

If there are several items that may leak or splash aggressive liquids, they can put in an area enclosed by a pony wall and a specific curtain. This is again is captured by a note on the P&ID.

18.2.5 Energy Hazards

Two types of energy hazards that may have a significant appearance on P&IDs are noise hazards and burning hazards caused by contact with hot surfaces.

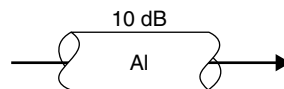
The strategies to put barriers for these two hazards are discussed here.

18.2.5.1 Noise Barrier

If the sound of a certain piece of equipment is higher than the limit set by the local regulatory body, action should be taken to meet the requirements of the local occupational health authority.

First of all it needs to be checked whether a certain noise is harmful or not. Generally speaking if you are in a noisy space and you need to shout to talk to a person in your vicinity, the noise is harmful for your ears.

Standards generally call for workplaces with less than 85 dBA noise as a safe workplace. This 85 dBA is measured



Acoustic insulation

Figure 18.1 Soundproof insulation.

over eight hours and taking the average reading (TWA: time weighted average).

The noise control can be done by isolating the noisy equipment. Isolating the noise source can be done by sound proof insulation or acoustical insulation.

For noisy equipment, the acoustical insulation can be placed by constructing a sound proof shelter or cabinet. This can be captured by a note in the P&ID.

If it is intended to minimize the noisy flow, the pipe may need to be acoustically insulated. In Figure 18.1, an example of a sound proof symbol on the P&ID is shown. "AI" stands for "acoustical insulation" and 10 dB refers to the value of decrease in sound level.

18.2.5.2 Burning Prevention

The way we prevent burning of personnel if they become in contact with hot surfaces is thermal insulation.

However this insulation is only for the purpose of burning prevention therefore it is named: "personal protection insulation."

In a personal protection (PP) arrangement, a thin layer of insulation around pipes, equipment, and containers is installed.

The purpose of "PP insulation" is only to prevent burning the operator when their hands or body come into contact with the hot equipment for as short as one second.

A metallic wall with a temperature above 60–70 °C is generally considered a burning surface. Different companies have different procedures for using insulation for PP. Some of them mention in their process guidelines that insulation must be used if the temperature of the pipe or equipment is higher than 60 °C, while some other companies use 65 °C or even 70 °C as the criterion.

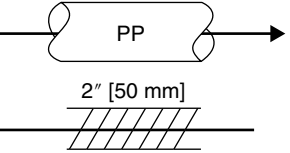
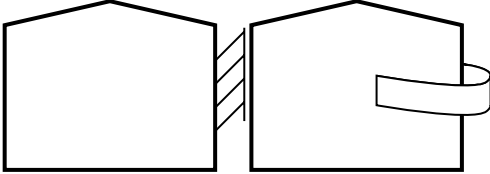
However, not all high-temperature equipment and/or pipes need PP insulation. If they are in remote areas or in inaccessible locations, PP insulation can be avoided if the company's guidelines allow it.

Personal protection insulation for pipes and equipment is needed if the temperature is a burning temperature and if operators could be in the vicinity.

Table 18.2 shows P&ID presentation of PP insulation for pipes, equipment and instruments.

The insulation thickness for PP is about 1–2 in.

Table 18.2 P&ID presentation of personal protection insulation for different items.

Pipe	Equipment	Instrument
	 <p>And/or in equipment call-out</p>	Not common

18.2.6 Safety Showers and Eye Washers

Safety showers and eye washers should be installed in different locations of a process plant. They could be on the ground, elevated, indoor or outdoor.

The purpose of safety showers and eye washers is to provide a quick and reliable source of water for an operator impacted by aggressive fluids. In the case of an accident, for example a splash of burning fluid like hydrochloric acid, the operator can run quickly to the closest safety shower/eye washer to wash his body and his eyes to minimize the consequences of the accident.

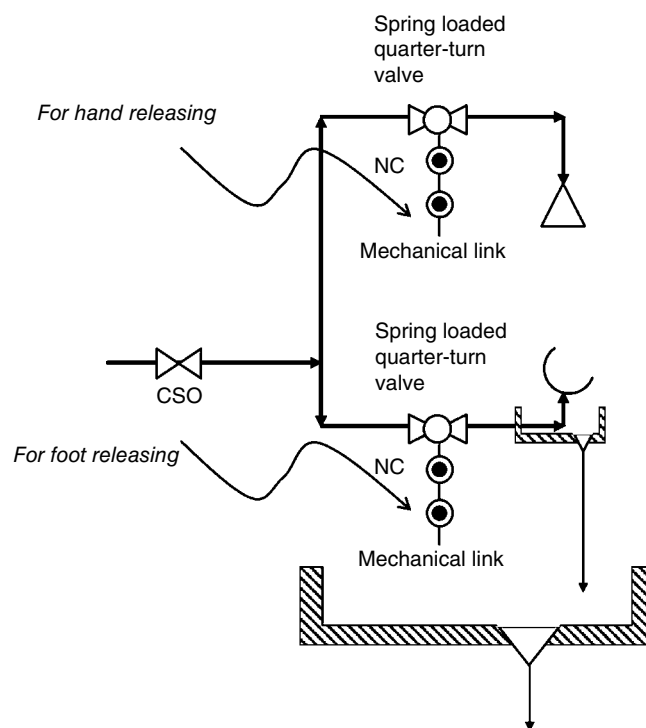
The decision for locating safety showers/eye washers is based on the aggressiveness of the handled fluid for operators. The aggressiveness of fluid can be found in the MSDS (material safety data sheet) of the handled material. If the MSDS states that a fluid is only toxic for a skin, a safety shower is enough but if it is toxic for skin and body, a combined safety shower and eye washer is needed. In some plants they put safety showers, eye washers, and combined safety showers plus eye washers in different locations depending on the MSDS of the handled material. However, in some other plants, for the purpose of standardization they put only combined safety showers and eye washers whenever it needs to install either of them.

Locating safety showers/eye washers only covers a specific radius around them and this radius is the maximum allowed running distance, which could be provided by the HSE (health, safety, environment department). The safety showers/eye washers could be placed on the ground and/or on elevated platforms depending of the location of the aggressive materials.

The water source for safety showers/eye washers could be potable water but in some plants they use the less preferable option, which is utility water.

The footprint of safety showers and eye washers are in two locations: the safety shower auxiliary drawing and potable water distribution network.

The P&ID of safety showers/eye washers could be part of auxiliary drawings. As safety showers/eye washers are generally the same within a plant a typical drawing could

**Figure 18.2** Typical details of a combined safety shower and eye washer.

be enough to show the details of all safety showers/eye washers within the plant.

Figure 18.2 shows the typical detail of a combined safety shower and eye washer.

The designer needs to check that there are enough provisions to make sure safety showers/eye washers always functional. For example if they are outdoors and the weather is very cold the designer may need to heat trace them. It is important to know that during the operation of a process plant all the safety showers/eye washers are tested periodically for their functionalities.

The other thing that the designer needs to consider is providing a sewer system or any other type of water receiving system for the contaminated water out of safety showers/eye washers.

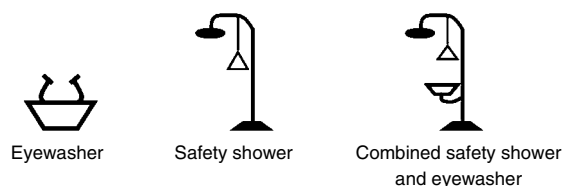


Figure 18.3 P&ID symbols for personnel emergency washers.

On the potable water distribution network a small symbol of safety shower or eye washer is enough (Figure 18.3).

There would be another drawing based on the plot plan to show the location of safety showers and eye washers in a plant.

18.3 Dealing with Environment

A plant is located in an environment. Therefore it is exposed to the changes of the environment's parameter.

There are plenty of parameters associated with environment including atmospheric pressure, ambient temperature, relative humidity, precipitation, radiation, dust, etc.

First of all, for obvious reasons, it is tried to establish a plant in the areas that have the least environment parameter harshness and swing.

If a plant, or a portion of a plant, should be established in a “bad” area and it cannot tolerate the environment parameter swinging, they can be “isolated” by locating them inside of a building. Indoor plants are very expensive and they are not always justifiable. However it is not rare to see fully indoor plants of food or pharmaceutical industries.

Table 18.3 summarizes the impact of several environmental parameters on process plants.

It can be seen that if the temperature swings are high this affects every single item in a plant. Heat transfer from the process to surroundings (or reverse) is not desirable and should be minimized. This is because heat transfer means a waste of energy. Therefore “isolating” the plant from the ambient temperature needs more thought.

To study the effect of temperature change on a process plant it would be better to study it in each stage of operation. The operations of a process plant can be normal and nonroutine operations. Nonroutine operations could be reduced capacity, upset, shutdown, and start-up operations.

Isolating a plant from environment in upset operation is no different from any other nonroutine operations and won't be discussed separately.

The requirements of start-up operation are generally considered in shutdown operation to make sure that after each shutdown, the unit or plant can be started up without any hassle.

In this context shutdown could be divided into short, medium, or long term shutdown. The different requirements of plants to be thermally isolated from the environment are shown in Table 18.4.

Table 18.4 shows that thermal insulation and tracing are used to minimize “communication” between the plant and its surroundings. In the next section we will explain the two main types of “thermal isolation”: heat conservation and winterization.

18.3.1 Arrangements for Maintaining the Temperature of the Process

One type of insulation is the type where the goal is keeping fluids at the temperature they are intended to be.

Table 18.3 Impact of environmental parameters on process plants.

Environmental parameter	Magnitude of swing	Process impact	Stack up strategy
Atmospheric pressure	Small swing	Units using ambient air (e.g. blowers)	Considering it in process design
Ambient temperature	Large swing	All equipment: heating up, aerial coolers	Considering it in process design (summer/winter designs)
Relative humidity	Large swing	Units using ambient air (e.g. blowers, burners)	Using dehumidifier or considering it in process design and/or P&ID development
Precipitation	Non-marginal	Open-top equipment (e.g. open-top tanks and basins)	Considering it in process design (e.g. closed-top tanks)
Sun radiation	Non-marginal but effective	All equipment: heating up	Considering it in process design (PSV for blocked fluids under sun radiation)
Dust	Depends on the location	Ambient air using units, open-top equipment	Using filter

Table 18.4 Thermal isolation of the plant from the environment.

Phases of plant operation	Risk caused by ambient temperature	Required action by the designer
Normal operation	There is a chance of temperature drop where it is undesirable. There is a low chance of winter problems for moving fluids in voluminous process items	It may need “heat conservation” insulation
Nonroutine operation (e.g. low production)	Chance of winter problems. Always low flow or no flow may causes winter problems	It may need “winterization” insulation and tracing
Short term shutdown (e.g. electricity outage)	High chance of winter problems. Always low flow or no flow may causes winter problems	It may need “winterization” insulation and tracing
Medium term shutdown (e.g. shutdown for maintenance, spare items)	Very high chance of winter problems. Always low flow or no flow may causes winter problems	It may need “winterization” insulation and tracing
Long term shutdown (e.g. shutting down plant because of lack of raw material or for economical reasons)	There is a high chance of winter problems and also long term corrosion.	Specific requirements are needed similar to filling items with anti-freeze, anti-corrosive liquids.

This type of insulation is called “heat conservation insulation” (HC insulation). HC insulation has a thickness that is generally larger than that of PP insulation. In HC insulation, a layer of insulation around pipes, equipment and containers is installed.

This insulation is for keeping the process at its optimum temperature and for the purpose of optimum plant operation.

As an example, if a liquid flows from container A to equipment B through a long pipe and the temperature of the liquid in container A is 80 °C and equipment B is designed based on a normal liquid temperature of 80 °C, then this means that no temperature drop along the connecting pipe between container A and equipment B is permissible (Figure 18.4).

If the pipe is long or the ambient temperature is very low, there could be chance of dropping the temperature of the stream far below 80 °C and therefore installing insulation is a must.

As a rule of thumb, heat conservation insulation is needed if the temperature of a flowing fluid is higher than the ambient temperature and also for process items that are small. The equipment in intermittent service may also need heat conservation insulation.

**Figure 18.4** Temperature drop when a hot service pipe goes through a cold space.

The thickness of insulation depends on several parameters including fluid (maintaining) temperature, ambient temperature, pipe or equipment size, wind velocity.

The ball park value of insulation thickness in inches is 1/100 of the fluid temperature of the pipe or equipment in degrees Celsius. This mean if piping has a service temperature of 400 °C, the insulation should be about 4 in. However, companies have tables in their guidelines to show the proper insulation thickness size depending on the pipe size and the service temperature.

The same problem exists when we are trying to transfer a cold fluid in high ambient temperatures, for example transferring cryogenic fluids in tropical countries. In such situations, again we can use insulation for process purposes, but its name is different; in this case it is called “cold conservation insulation” (CC insulation).

As a rule of thumb cold conservation insulation is needed if the temperature of the flowing fluid is less than the ambient temperature or less than the dew point of ambient air.

Table 18.5 shows the P&ID presentation of heat conservation insulation for pipes, equipment and instruments.

It is the judgment of the process engineer to decide about the need for HC insulation for each item in a plant. There are, however, some generalizations that can be used (Table 18.6).

18.3.2 Winterization

The colloquial concept of winterization is preparing a system for winter to protect it from harsh conditions to make sure it will be quickly functional after the winter.

Table 18.5 P&ID presentation of heat conservation insulation for different items.


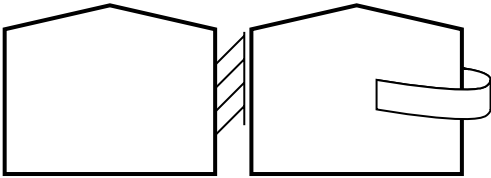
Pipe	Equipment	Instrument
 And/or in pipe tag 114-ASL-COS-100-38H EHS-07171-2-B9W-50H	 And/or in equipment call-out	Not common

Table 18.6 Pipes that most likely do not need HC insulation.

- Pipes go to ponds
- Pipes go to coolers (cooling stream of heat exchangers)
- Pipes of “used” cold streams (e.g. CWR)
- Short and large diameter pipes

In the process plant world, winterization basically means implementing specific features in a plant design and P&ID development to prevent the impact of cold weather on plants in shutdown conditions.

Winterization also can refer to activities to make a piece of equipment functional even in the harsh conditions of winter.

Therefore in a broader sense, winterization is activities to prevent freezing, frosting, or setting of matter in a process plant in its all operation phases.

The other name of winterization is “Frost prevention” or “freeze protection.”

When a plant is shut down, either partially or completely, after de-energizing the plant elements, the next step is to drain all the pipes, equipment and containers to make sure there is no trapped liquid in enclosed spaces, and that all of the enclosures are empty. This is to protect the plant during post-shutdown time to keep the plant safe against anomalies caused by trapped liquids. Trapped water in a plant may freeze and expand, and this expansion of frozen water can rupture pipes, equipment or containers. Trapping very heavy oil in plant enclosures will cause it to set and become hard to move. After a long shutdown, this makes the plant’s start-up very difficult.

For all of these reasons, all the trapped liquid should be emptied by the plant operator. However, the problem is that each plant may have a few hundred or more drain valves and only a small number of operators. Therefore, draining all the trapped liquid through a few hundred drain valves may take weeks to complete and during this

time if the ambient temperature reaches a low level (and the word “winterization” comes from here), it may endanger the plant’s equipment.

There are passive winterization protection and active winterization protection methods.

The passive winterization protection methods are generally more inexpensive than active methods. The passive methods could be implemented instead of or in addition to active methods.

The winterization methods are listed in Table 18.7.

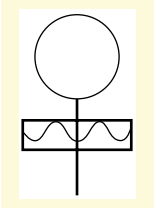
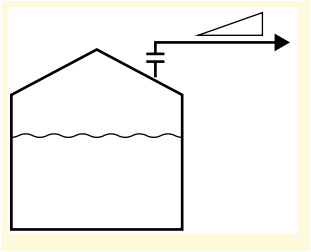
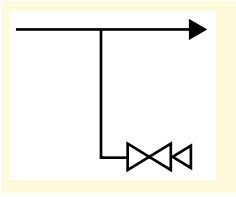
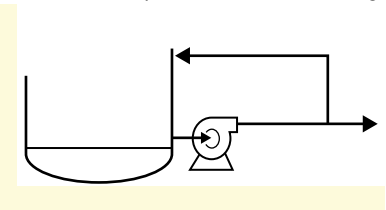
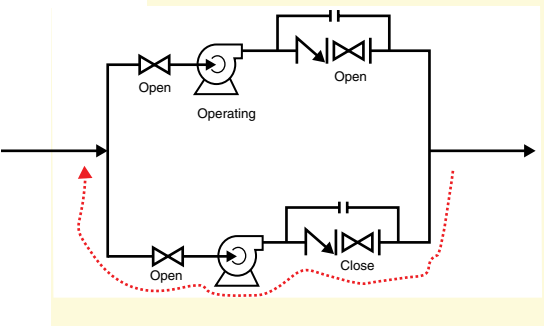
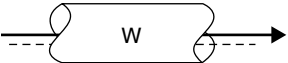
One passive method of dealing with winterization is minimizing the exposed area of process items. Such minimized exposure can be obtained by putting process items indoors, inside buildings or inside sheds/cabinets or by burying underground pipes or containers below the “frost line.” The frost line is an imaginary surface below the ground surface under which the soil is less affected by the atmospheric temperature. It is assumed that the (wet) soil below the frost line does not freeze in the winter.

For fluid-in instruments like Bourdon tube pressure gauges, winterization can be attained by essentially not sending service fluid inside of the instruments. When freezing is an issue in a specific environment and a specific service fluid, specific pressure gauges with filled liquid can be used. In such Bourdon tube pressure gauges the Burdon tube is filled with a non-freezing liquid (like glycol) and capped with an elastic membrane. The membrane (diaphragm) allows pressure to transmit to the Bourdon tube without allowing the problematic service fluid getting in to the Bourdon tube.

The other good practice regarding dealing with winterization is providing sloped piping toward “tolerable” equipment. This technique is known as providing “internal natural free draining.” An example of more tolerable items against freezing are tanks.

Elimination of dead legs is important if implementing the concept of “natural internal drainage.” Dead legs are

Table 18.7 Winterization methods.

	Examples
Minimizing exposed area	<ul style="list-style-type: none"> Indoor equipment Underground and below frost line equipment Diaphragm seal (for fluid-in instruments) 
Internal free draining	
No dead legs	
Mechanical-driven flow	<ul style="list-style-type: none"> Recirculation by a fluid mover on emergency power  <ul style="list-style-type: none"> Warm up system for spare item 
Thermal-driven flow (thermosiphon circulation) Winterization insulation and tracing	

problematic from different aspects and should be eliminated, or at least minimized. From a winterization prevention viewpoint, dead legs are the source of the start and promotion of freezing. Dead legs can be eliminated by minimizing the length of dead end pipes, pipes that are blinded or have a normally closed valve on them.

A fully flowing stream doesn't freeze. The flow can be initiated mechanically or thermally.

The mechanical flow could be generated by a pump on emergency power to ensure the stream is always flowing. The flow can be generated through a spare pump by its mate operating pump.

Thermally driven flows are known as thermosiphon circulations. A source of energy can generate thermosiphon circulation in a non-obstacle space inside of process items.

The other technique of winterization is heat tracing covered by insulation to prevent freezing or setting of non-drained/trapped fluids.

In some cases, where the viscosity of the trapped liquid is higher than 4000–5000 cP, pipe jacketing is used instead of pipe tracing (Figure 18.5).

Heat tracing and thermal insulation is, again, for process reasons but unlike HC insulation (or CC insulation), it is not for process purposes during normal plant operation. Instead, it is to minimize the problems that may arise after a plant shutdown and/or when it is decided to start the plant back up.

This type of arrangement is called “winterizing” (W).

In a “W arrangement,” pipes, equipment and containers are brought into contact with a few tracer lines and then the lines are covered by insulation.

The tracer line could be a tube of steam, a tube of hot glycol or any other hot fluid, or “electrical tracing.”

Figure 18.6 shows a winterized pipe before being wrapped with insulation. In this application two tracer tubes are used but it could be higher or lower number of tracer tubes.

Electrical tracing is basically a resistance element that will be hot when it carries electrical current.

On the P&ID three features of winterization heat tracing and insulation should be shown: symbol, type of tracer, and holding temperature. In some companies the thickness of insulation is shown too.

Figure 18.7 shows two examples of winterization heat tracing and insulation.

For each winterization system, a “holding temperature” should be specified. The holding temperature is the temperature that should be held by the winterization



Figure 18.5 Two main methods of active heat conservation.

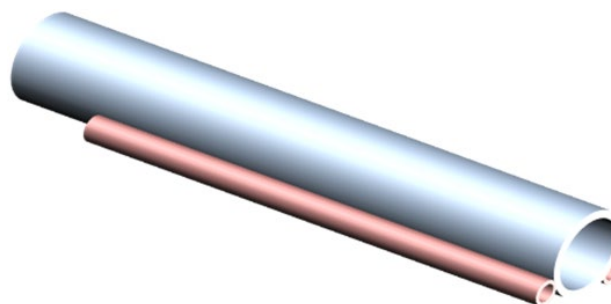


Figure 18.6 Heat trace arrangement.

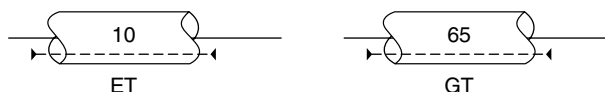


Figure 18.7 P&ID representation of steam tracing, glycol tracing and electrical tracing.

arrangement to protect the system from the dangerous effects of “trapped liquids.” The holding temperature is usually 5–10 °C higher than the “problematic temperature” for the trapped liquids. The problematic temperature could be the liquid's freezing point or its pour point, or in the case of a gas stream, its dew point.

Based on the required holding temperature for winterization insulation and tracing, the designer (generally the piping designer but sometimes the process designer) determines the size and number of the tracer tubes.

There are a few cases where a winterization arrangement (insulation plus tracing) is used not for the purposes of winterization but rather for process temperature conservation. For example, sometimes insulation plus a tracer is installed on the vapor pipes from the roof of a tank to prevent condensation of vapors, which causes liquid droplets, which in turn may cause corrosion in the pipe. However, we still call this system a winterization system.

Each of the above-mentioned insulation and tracing arrangements has a symbol on P&IDs, which are shown in Table 18.8.

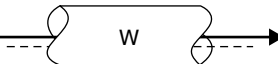
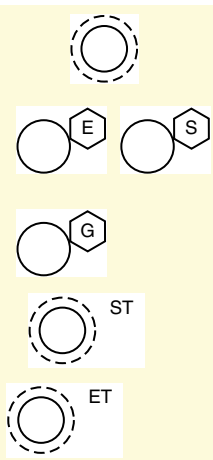
The decision on the requirement of winterization is easier than instruments. For instruments we need to go a bit further into the structure of it to be able to check if the instrument needs or doesn't need winterization provisions.

Here are some rules which help us for making decision.

First of all, instruments that are not in contact with process fluid don't need winterization provisions.

If the instrument is of inline type (like control valves and the majority of flow meters) it is handled together

Table 18.8 P&ID presentation of winterization for different items.

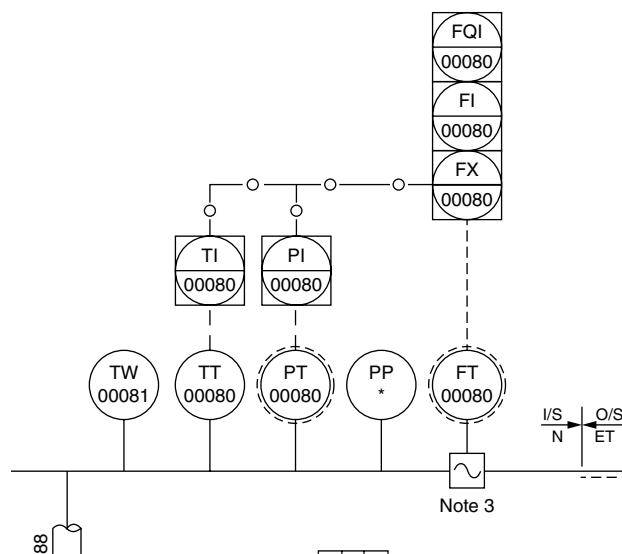
Pipe	Equipment	Instrument
 And/or in pipe tag 6'-HLS-AS-1003-64H-GT P-07111-2-B39W-50H-ET 222-163L-2"-lhST(w)(1-1/2")	Not common unless for small equipment	

with the pipe it is installed in. If the pipe needs winterization provisions, the inline instrument also needs it.

For non-inline (offline) the requirement depends on whether it is fluid-in type or not.

The fluid-in instruments like Bourdon tube in pressure gauges may need winterization arrangement. But non-fluid-in types definitely don't need any winterization arrangement.

Figure 18.8 shows some examples of winterization of instruments.

**Figure 18.8** Winterization of some instruments.

18.3.3 Deciding on the Extent of Insulation

From a purely theoretical viewpoint, when it is decided that an “item” is to be insulated, the “whole” item should be insulated. However in reality we don't always insulate whole the item, unless it is fully justifiable from an economical viewpoint. The full insulation may also make the normal operation of the system of interest problematic.

There are cases where the designer needs to decide on the extent of insulation. This means he needs to decide if all the pipes/equipment of interest should be insulated or only a portion of them.

The extent of insulation is generally mentioned on P&IDs; on the main body of the P&ID or in the note area.

For example, one general question is whether a full tank, including body and roof, should be heat insulated if the tank needs to be insulated or not. All the tank could be insulated if it is decided so, but the cost of insulation could be saved by insulating the body of the tank. As the main reason for this insulation is to keep the liquid content “warm,” possibly the roof of the tank doesn't need to be insulated. Even on the body of the tank, only the portion that is in contact with liquid could be decided to be insulated. The additional reason for not insulating the roof is that the heat transfer coefficient of gas/vapors is much less than for liquids, and the heat transfer from the roof is already low. A company may decide that: “if a tank needs to be heat insulated, only the body of the tank up to the high liquid level should be insulated.”

For pipes almost always all the body of the pipe could be insulated. The only exception could be large bore

pipes with partial flow. In such cases if the budget is not enough, only the lower portion of the pipe (up to a certain angle) could be insulated.

For valves there are different types of insulation extent. A valve could be insulated completely or partially. The complete insulation of a valve means insulating its body and its bonnet. As the main bulk of fluid is inside of the “body” of a valve the main heat insulation happens by insulating only the body of the valve. However, if the completeness of insulation is very important, it could be decided to insulate the bonnet of the valve too. The extent of valve insulation could be mentioned as an acronym beside or below valves as “B” (for body insulation) or “B&B” (for body and bonnet insulation). Obviously a valve that has insulation on its bonnet is more difficult for inspection and repair.

Where there are heat tracers below the insulation it is easier to decide on the extent of insulation; as we have expended money on heat tracing, it is better to keep the equipment as insulated as possible.

Table 18.9 lists priority parts for insulation for several types of equipment.

Table 18.9 Insulation priority of items.

	Coverage of insulation (if insulation needed)	Priorities
Pipes	Whole pipe but left out the flanges	Not applicable
Valves	Partially	Priority 1: body Priority 2: bonnet
Vessels	Whole vessel	
Tanks	Partially	Priority 1: wall Priority 2: roof
Pumps	Whole pump	Not applicable
Compressors	Whole compressor	Not applicable
Drivers	Electric motors: no, they generally need cooling Steam turbine: full insulation	Not applicable
Heat exchangers	Shell and tube: full insulation Plate: generally no insulation Spiral plate: full insulation Aerial cooler: No insulation	Not applicable
Safety devices	Partially	Priority 1: inlet and outlet pipes Priority 2: PSV body (but not rupture disk)

Table 18.10 Different requirements in insulation systems.

	Insulation	Heat tracing	Other considerations
Heat conservation	Yes	No	No
Personal protection	Yes (thin)	No	No
Winterization	Yes	Yes	Yes

18.3.4 Summary of Insulation

Table 18.10 shows the summary of requirements in different insulation systems.

18.4 Utility Stations

Utility stations (US) are pretty similar to potable water fountains that you may see in malls and other public areas. The difference is that US provides utilities more than only potable water.

USs are boxes which contain different utilities for process plants during inspection, maintenance, and repair. They may also be used during normal operation but in short period of times.

Whenever an operator needs to use any of these utilities, he needs to connect a piece of hose to the connection inside of the US and then open its valve.

The utility in each US could be a steam, air, water, nitrogen gas, electricity, etc.

All these utilities, except electricity, are named “process utilities.”

A schematic of a utility station is shown in Figure 18.9.

Figure 18.10 shows two different ways of showing utility stations on P&IDs.

On the left-hand side of Figure 18.10 the utility streams in USs are plant air (A), steam (S), and plant water (W). On the right-hand side of Figure 18.10 a US with an additional utility stream of nitrogen gas (N) is shown.

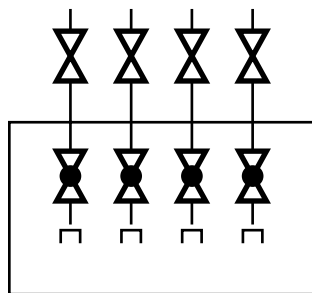


Figure 18.9 A utility station on a P&ID.

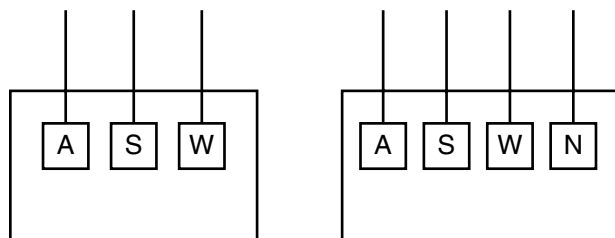


Figure 18.10 P&ID presentation of utility stations.

In the text below we discuss the usage of each of these utilities.

Steam: a steam that is used in USs is named utility steam. We need utility steam whenever we need to clean stubborn fouling and scales. Generally speaking, steam is the last resort for removing dirt from process equipment. If utility steam fails to clean fouled or scaled equipment, the next available option is cleaning with solvents or chemicals, which is a very expensive operation.

Air: utility air or plant air is again used to clean dirty equipment. Utility air is more effective on dry dirt. The other usage of utility air is driving power tools. The quality of utility air is less than the quality of instrument air. However it should be clean and free of any dirt.

Water: utility water is the water from the US. This water could have quality similar to potable water but in the majority of cases its quality is less than potable water. So it is important to know that utility water is not always drinkable. This water shouldn't have any scaling or fouling tendency. Utility water again is used for washing and cleaning of dirty equipment.

Nitrogen gas: nitrogen gas could be implemented in USs for different reasons. The main property of nitrogen

gas is its inertness. Thus nitrogen gas can be used for purging and pushing toxic, aggressive, flammable gas from an enclosed space. This action could be taken before letting people enter inside a piece of equipment like a vessel or a tank, or nitrogen gas can be introduced in a process system before starting up.

The type and number of utilities in each utility station may vary depending on the equipment in the vicinity. It is important to note that the hoses that could be used for a US have a standard length. Therefore each US can only cover a specific radius around itself. The designer needs to decide the location of the US and the utility in each US. The systematic way to do this is by preparing a list of equipment in the plant and then listing all the required utilities for each piece of equipment during maintenance. Then, based on a plot plan, the designer can decide upon the number of required USs in the plant in a way that covers all the equipment (Figure 18.11).

Utility stations are not necessarily located on the ground floor and they could be on the platforms around the equipment too. However, elevated USs generally have only gas or vapor utilities. If a utility like "plant water" is to be provided for an elevated utility station, provisions should be considered to prevent splashing plant water on the personnel who are unaware and passing by on the ground.

In each plant, before developing US drawings, the plant owners should be consulted for the availability of different utility steams, their pressures, and the length of the standard hose for USs.

The exact location of a US cannot be recognized from P&IDs. However, a US can be seen in utility distribution drawings (Figure 18.12).

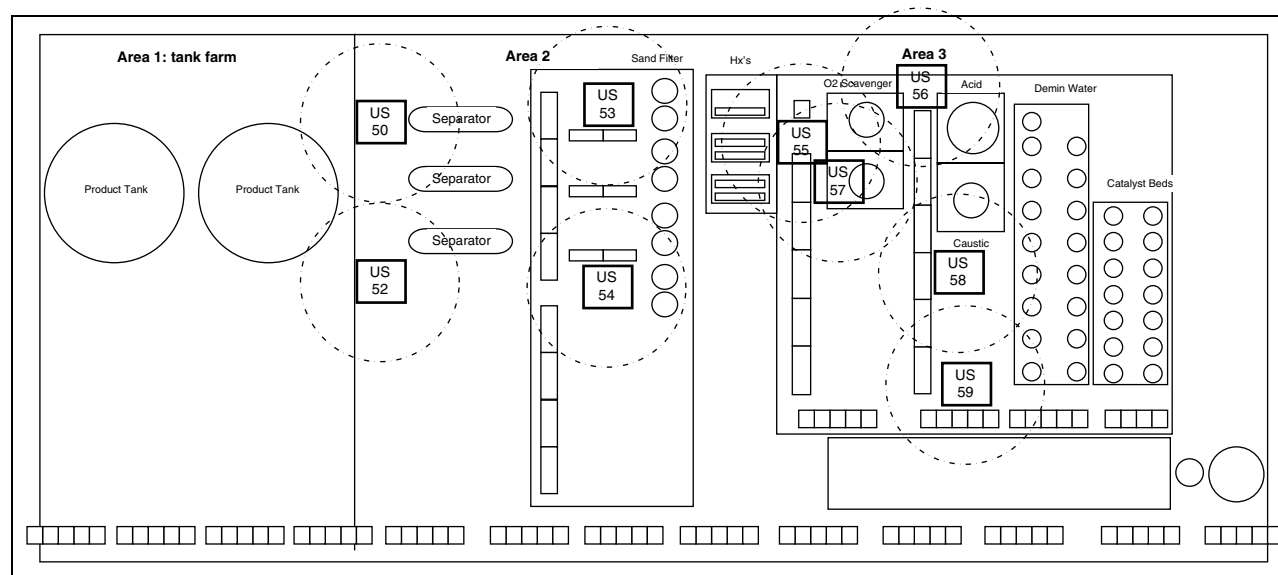


Figure 18.11 Positioning of utility stations in a plant based on the plot plan.

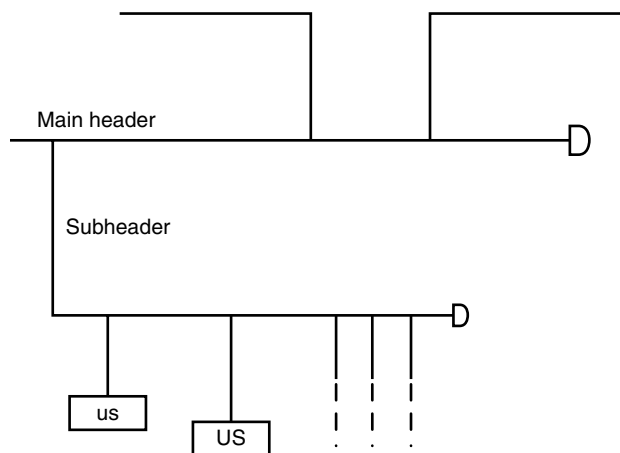


Figure 18.12 A utility network connected to a utility station.

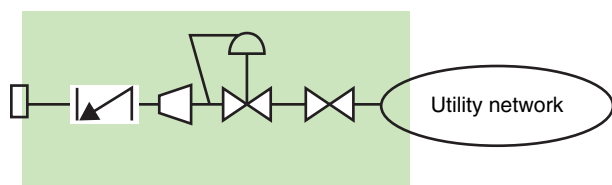


Figure 18.13 Detail of utility streams in utility stations.

Even though, depending on the type of equipment in each area, the utility streams in each utility station could be different, some companies decide to install just a standard US in which there are a fixed number and type of utilities in all of them. They have made the decision to do this so as to not confuse operators; they will know whether a specific US does or doesn't have, for example, utility steam.

Utility stream pipes are generally 2" or smaller than that.

The detail of each utility stream pipe up to the point for usage by the operator could be different depending on the type of utility stream. However the arrangement of isolation valve, pressure regulator, check valve, and hose connection is very common (Figure 18.13).

The isolation valve is needed to isolate the downstream if there is an issue. The isolation system could be decided to be more complicated than a single isolation valve – e.g. double block and bleed – in some cases.

A regulator could be needed to adjust the pressure to a pressure that is not harmful for operators.

A check valve can also be placed at the last point of a utility pipe before connection to process to make sure contamination of utility fluid by process fluid is prevented.

It is important to note that, as all the USs are connected to a utility network, the pressure of utilities at the edge of each US may fluctuate during the usage of the utility. For example, if three USs that are located very close to each other using utility water at the same time, the utility water

pressure will definitely be less than the pressure of utility water if only one US is functioning. Because operators don't want to see any pressure fluctuations in any utility steam, they will generally use a pressure regulator to adjust the pressure in the utility network to a fairly constant and non-harmful pressure for the operator.

18.5 Off-Line Monitoring Programs

We need off-line monitoring programs wherever we need to check the "process properties" of a stream but there is no process analyzer available. The word "available" here is used for the situations where there is no process analyzer in the market to measure the process parameter of interest, or where the cost of the available process analyzer is beyond the plant budget, or the criticality of the parameter is so low that doesn't justify the purchase of a process analyzer. In such situations if (and this is an important if) the process parameter is sluggish enough, an off-line monitoring program can be used.

If the process parameter is very agile, which means the process parameter changes a lot and very quickly, off-line monitoring programs don't work.

18.5.1 The Program Component

An off-line monitoring program is defined as: taking a sample, protocol to transfer the sample to the lab, suitable testing procedure to measure the process parameter, and sending the result to the plant operator to take the appropriate action.

Therefore, an off-line monitoring program works in a similar way to a control loop and the difference is that in an off-line monitoring program some automatic actions are replaced with human actions. This concept is depicted in Figure 18.14.

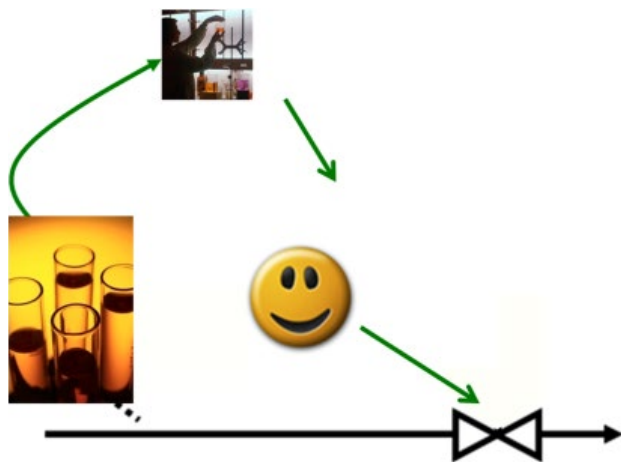


Figure 18.14 Off-line monitoring programs.

The only footprint of a sampling program on the P&ID is the sampling system. Therefore our discussion is limited to the sampling system.

18.5.2 Sampling System

A sampling system is a set of hardware (system) to extract the sample from the process and handing-over to the lab through the hands of a process operator.

A sampling system can also be used to provide samples for a process analyzer.

The type of sampling system used depends on different parameters. The parameters that affect the type of sampling system are: the phase of sample (liquid, gas/vapor, flowable solid), the nature of the sample (temperature, pressure, innocent or non-innocent), the frequency of the sampling, dead leg versus loop sampling, the policy of handling residual samples, etc.

A sampling system for liquid is different from that for gas, and it could be different from those for vapor or a flowable solid.

The sampling system could be more complicated if the frequency of sampling is needed to be higher. If a sample should be taken very frequently (possibly because the process parameter is not sluggish enough) it could be an automatic sampling system. In an automatic sampling system, samples are taken in a specific, predetermined intervals (grab samples) and they can be sent to lab one by one by the process operator or mixed together and sent in a larger volume (composite sample) to the lab by the operator.

A sampling system could be based on a dead leg system or a loop system. In a loop system as soon as the sampling valve is opened, the representative fluid is discharged. While in dead leg sampling the fluid remaining in the sampling tube from the previous sampling should be emptied, to have a fresh, representative sample.

Each sampling system includes six main portions:

- 1) Sample extraction device
- 2) Sample transferring tube
- 3) Sample conditioning system
- 4) Sample hand-over system
- 5) Waste sample collection system
- 6) Sampling station structural frame.

Figure 18.15 shows a block diagram of a sampling system.

A sampling system is more expensive that it seems. With more need for having accurate data and accurate



Figure 18.15 Block flow diagram of a sampling system.

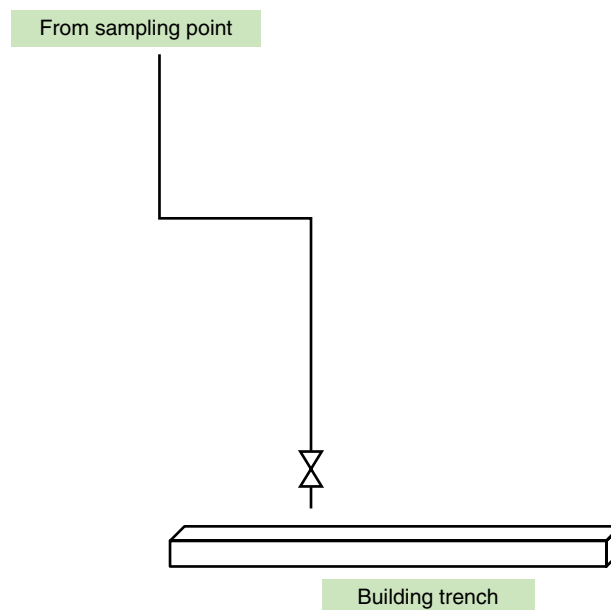


Figure 18.16 Simplest type of sampling system.

sampling, and also ever tightening environmental regulations, sampling systems are getting more and more expensive. There are different ways to bring down the cost of a sampling system to a more reasonable cost. One way is standardization of sampling systems within a plant and the other way is developing common sampling stations for several sample points rather than dedicated sample station for each single sample point.

A sampling system could be as simple as a single tube with single valve on it above a trench (Figure 18.16). However, it could be as extravagant as a large cabinet with multiple pieces of equipment in it (Figure 18.17).

In the sections below these six components of sampling systems are discussed.

18.5.3 Sample Extraction Device

To obtain a representative sample it should be taken from suitable location by a suitable device.

The most accurate sampling device to extract samples from fluid pipes are “sampling quills.”

A sampling quill is a device that extracts samples from the center of pipes and functions based on an isokinetic concept. However, there are cases where a simpler sample extraction device is less expensive but still acceptable from a technical view point. In some plants, a

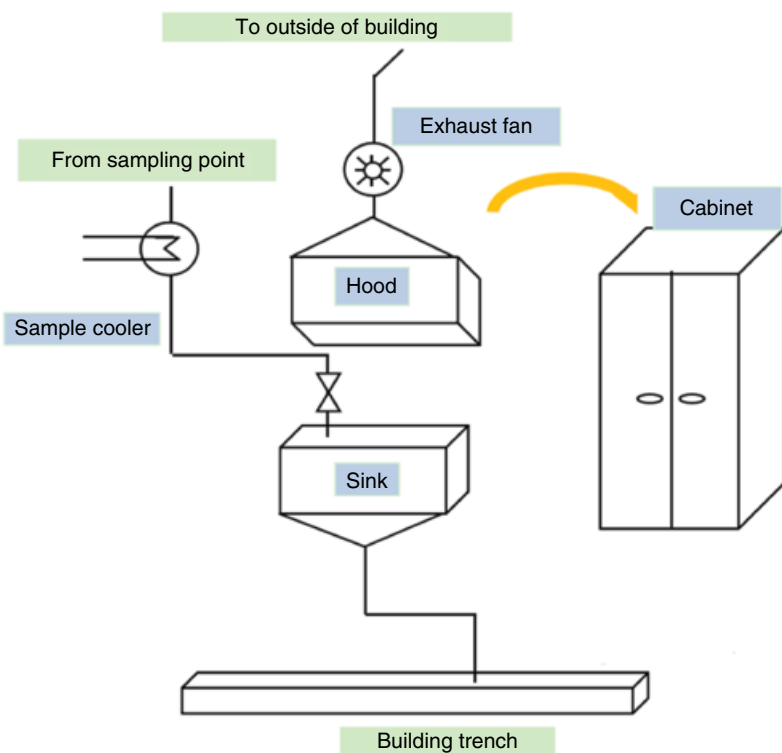


Figure 18.17 A luxury sampling system.

simple drain valve is used for sampling purposes, which is not the best decision.

On P&IDs, sampling quills can be tagged as “specialty items.”

18.5.4 Sample Transferring Tube

The goal of a sample transferring tube is conveying the sample from a sampling point to a sample hand-over point. These two points are not necessarily close to each other because each has their own set of selection criteria.

The sample point should be a point which is suitable from process and non-process viewpoints.

The sample point from process viewpoint should be a point that provides a representative sample and has enough pressure to get to the sample hand-over point. The non-process criteria are safety and the convenience of the operator. A sample point is preferably from a process point with low pressure (e.g. suction side of pump rather than pump discharge side), low “inventory” (e.g. smaller vessels rather than large tanks), and low temperature points.

The best sample hand-over point is a point that an operator can conveniently, comfortably, and stably get the sample from the system.

A sample stream generally has a flow rate of about 1 l min^{-1} . The sampling conductor is generally a tube with a nominal size of less than 1" and commonly $\frac{1}{2}$ " or $\frac{3}{4}$ ".

The sample transfer tube should be kept at the smallest length to make sure the sample can get to the hand-over point with the existing original head. The tube length can be kept at minimum by placing the sample station as close as possible to the sample point.

If the sampling system is of dead leg type the diameter of the sample transfer tube should be kept at a minimum too; this is to minimize the lag time of the system and also to get an as fresh as possible sample to the hand-over point. After each sampling a portion of non-delivered sample remains in the sample transfer tube in a dead leg system. For the next sampling, it should be made sure that the remaining portion of the previous sample is emptied from the tube to make sure a representative sample is taken. A short and narrow sampling tube also decreases the purge/drain time.

There is probably no constraint on the tube diameter in a loop type sampling system.

18.5.5 Sample Conditioning System

As the sample goes to the “hands” of the operator, it should be safe for him. As a minimum it should be assured that the pressure and temperature of the sample is safe.

The pressure generally gets to a safe region after the sample passes through the narrow tube. However, if there is doubt, a pressure regulator could be installed on the sampling tube.

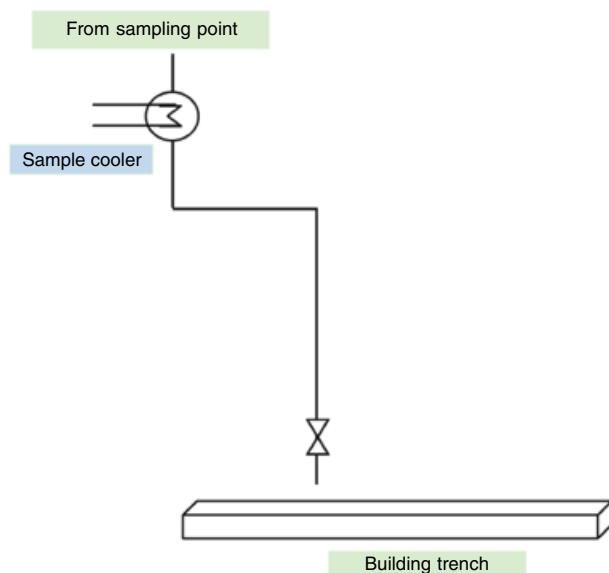


Figure 18.18 A sample cooler on a simple sampling system.

If the temperature of the sample is in a region that can be considered as “burning” (say more than 65 °C), a sample cooler should be placed to decrease the temperature to a safe temperature.

Figure 18.18 shows a sampling system with a sample cooler (with PFD symbol).

There have been attempts to make common units for conditioning multiple sample tubes, each multiple stream sample coolers.

18.5.6 Sample Hand-Over System

The sample hand-over arrangement could be simply a needle valve or ball valve.

If it is intended to design a foolproof arrangement to make sure no valve is left open in the plant, a spring type ball valve can be used to close off the valve automatically after opening.

18.5.7 Waste Sample Collection System

There could be residual sample wasted during the sampling operation. The sample release could be in the form of liquid spillage or gas/vapor release.

There is more chance of sample waste in dead leg sampling systems rather than loop sampling systems.

In older days, the waste sample was discharged to drain if it was liquid and to atmosphere if it was gas/vapor.

However, these days, with stricter environmental regulations and tighter budgets, plants have more tendency to recirculate the waste sample back to the process as much as possible.

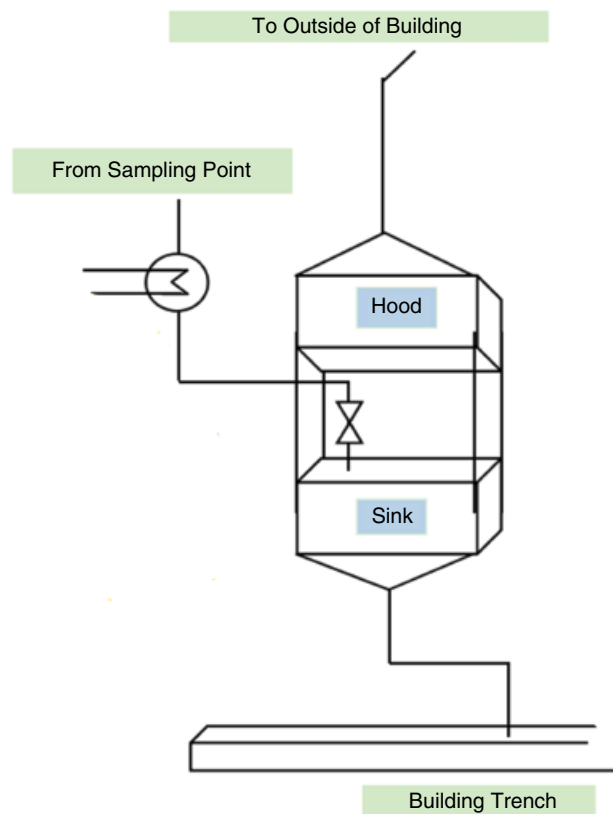


Figure 18.19 A sampling system with sink and hood.

To collect and recover the waste sample a sink and hood is needed to collect the waste liquid and waste gas/vapor respectively (Figure 18.19).

Generally speaking equipping sampling system with sink or hood or both of them is justifiable where there are several sampling tube to a sample station.

18.5.8 Sampling Station Structural Frame

The sampling station structure should be installed in locations within the plant in a safe and comfortable location for the operators. For example, it is a better idea to place them on the ground rather than on platforms or on the roof of the tanks. If there is a chance of a meeting aggressive vapors or a spill of aggressive liquids it should be taken care of in a sampling system.

A sampling system could be outdoors or indoors. It could be simply on a rack, or in an open box or door box (cabinet). The cabinet could be walk-in or non-walk-in type. In some cold areas, operators prefer to see all the sampling systems inside of heated walk-in cabinets. If a sampling system is inside of a walk-in cabinet it should be checked whether it is categorized as a “building” and if the relevant building code is applicable or not. The other issue regarding walk-in cabinets is whether they

should be ventilated for the safety and comfort of the operator or not.

18.5.9 Showing a Sampling System on P&IDs

There are different ways to show sampling systems in P&IDs. One way is to show the sampling system on each sampling point (Figure 18.20).

Another method is just to put in a reference flag on each sampling point and then refer to an auxiliary P&ID for the detail of the sampling system. The goal of this method is to develop a less congested P&ID.

In this method all the sampling stations can be moved to auxiliary drawings and in the main process drawings only a reference number is left (Figures 18.21 and 18.22).

Here “SC” stands for “sample connection.”

If this is the practice, the different sampling stations are categorized in several sampling groups (e.g. type 1, type 2, etc.) to make the referencing easier.

A typical sampling system may appear on an auxiliary drawing as shown in Figure 18.22.

18.5.10 Sampling System for Process Analyzers

A sampling system may be needed for non-flow through process analyzers. These are a type of automatic sampling system.

A process analyzer sampling system could be as pricy as the process analyzer by itself or even more expensive than that.

A sampling system for a process analyzer is mainly developed based on the requirements dictated by the process

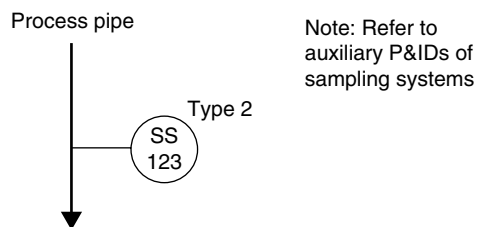


Figure 18.20 Sampling system on the main P&ID.

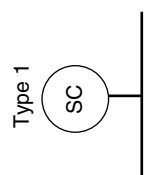


Figure 18.21 Sampling system flag on the main P&ID referring to auxiliary P&IDs.

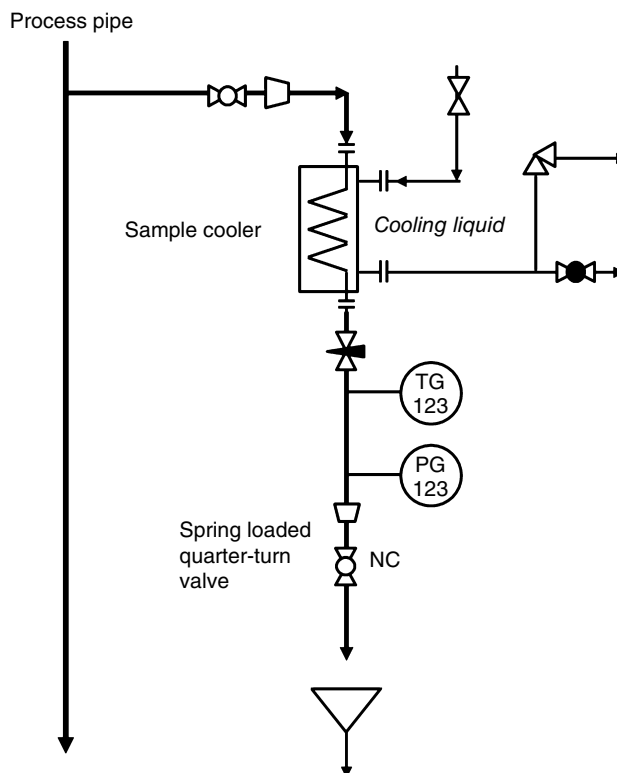


Figure 18.22 Sampling system on auxiliary P&IDs.

analyzer vendor. However, they are very different from manual sampling systems. They may have some elements of manual sampling systems. They may have sample extraction device, sample transfer tube, sample conditioning system, and a station containing all the mentioned elements in addition to the process analyzer.

Similar to a manual sampling system, it is not very preferable to show the detail of process analyzer sampling system on the main P&IDs. They may be shown only on the vendor P&ID or on an auxiliary P&ID.

18.6 Corrosion Monitoring Program

There are some cases that corrosion should be measured in one or several “critical” points of the plant. Such need could be for evaluating the success or lack of success of the anti-corrosion program in the plant or for other reasons. The corrosion rate parameter, which is the output of the corrosion monitoring program, could be used to adjust the corrosion inhibitor injection system and/or recording it as the lesson learned for material selection in future similar projects.

Such corrosion monitoring can be done on-line or off-line. An on-line system is in essence a process analyzer system, which was discussed in Chapter 13.

The off-line corrosion monitoring program is a type of off-line monitoring program, which was discussed in the previous section.

An off-line corrosion monitoring program is a set of hardware (system) and procedures to measure and report the “corrosion rate” in a specific location of a plant. A corrosion monitoring program specifies the required corrosion coupon system, procedures to transfer the coupon to the lab, applying a test procedure to measure the corrosion rate, and sending the results to the appropriate parties.

The only footprint of a corrosion monitoring program on the P&ID is the corrosion coupon. Therefore our discussion is limited to the corrosion coupon.

A “corrosion coupon” is a piece of material with the specific shape and specific weight. The material of the corrosion coupon is generally selected the same material of pipe or equipment. Corrosion coupons are located in locations to provide meaningful information. They are generally located at locations that are suspected as having a high corrosion rate, upstream of some critical equipment and mainly on pipes. The locations where the fluid is stagnant are not good candidates for a corrosion coupon.

They should also be located in accessible, safe, and comfortable locations for operators.

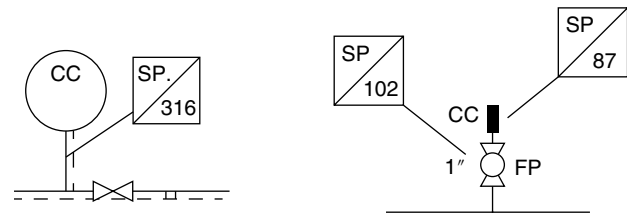


Figure 18.23 Corrosion coupon.

Figure 18.23 shows two schematics of a corrosion coupon on P&IDs.

18.7 Impact of the Plant Model on the P&ID

Generally speaking the plant model is developed based on the P&ID and other documents like the plot plan.

However, there are cases that the route is reversed, which means the P&ID needs to be changed because of the plant model (Figure 18.24).

Space constraints can dictate changes on the P&ID but not all of them are acceptable from a process viewpoint.

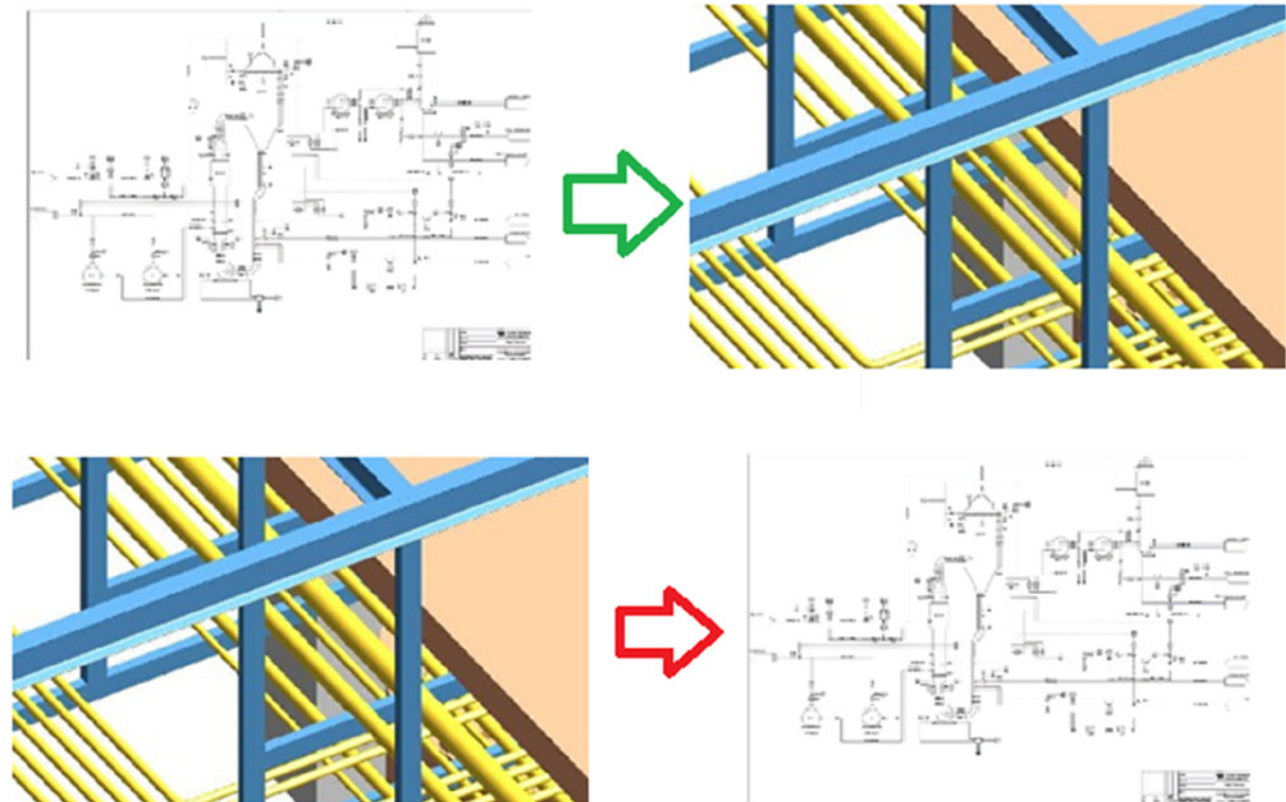


Figure 18.24 Direct route and reverse route.

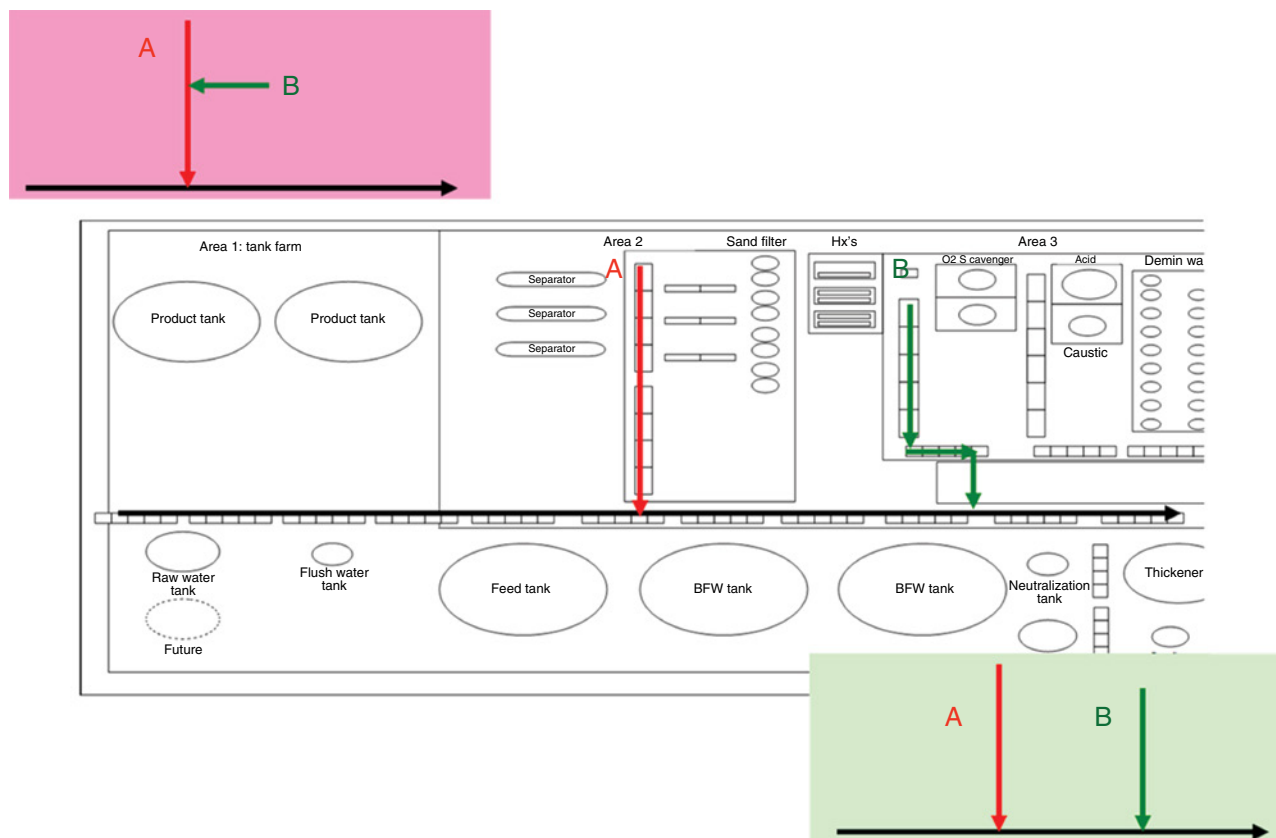


Figure 18.25 P&ID change because of piping model.

The reason for forcing changes in the P&ID because of the plant model is mainly space constraints. The change on P&IDs because of the plant model are mainly for items installed on pipes and not for large equipment.

The space constraints may push an item from indoors to outdoors (or vice versa), change on tie-points on pipe networks, etc.

Figure 18.25 shows such issue. On the P&ID it is shown a pipe “B” tying into pipe “A” and then the header (top schematic). Later during the development of plant model, it was realized it was not doable and pipes “A” and “B” should be tied into the header separately.

18.8 Design Pressure and Temperature Considerations

In process plants flow goes through pipes, containers, fluid movers, heat exchangers and other equipment. The temperature and pressure of flow may go up and down depending on the action of equipment on the flow and also the ambient temperature.

The responsibility of a designer is not only making sure a piece of equipment does its duty but also making sure

that it keeps its structural integrity in all conditions during the life cycle of a plant when pressure and temperature goes wildly high or low.

This assurance is not always checked during the design stage of equipment and piping. Therefore during P&ID development it is very critical to check the structural integrity of process items.

Generally speaking pressure and temperature are the most important parameters where their fluctuation may harm a piece of equipment. We explained in Chapter 12 why, out of five process parameters of flow rate, level, temperature, pressure, and composition, we only focus on pressure and temperature.

The structural integrity of process items are primarily defined by the design temperature and design pressure of the process item. Without going through the complicated mechanical engineering concepts, the design pressure is defined as the pressure where if the operating pressure stays at it (or over it) in the “long term,” the item will be ruptured.

The other definition is: “design pressure is the pressure that an enclosure can tolerate and not yield to on a long term basis.”

It is important to know that stating the “design pressure” without mentioning the “design temperature”

is meaningless. The “design pressure” should be mentioned “at” a “design temperature” as a pair; e.g. the design pressure of this vessel is 900 KPag at design temperature of 80 °C or “900 KPag @ 80 °C.”

For all process items a wise pair of “design pressure @ design temperature” should be selected and requested from the item vendor.

Moreover, this pair should be coincident. This means that during the operation of a process item a pressure as high as the selected design pressure could happen during the time the temperature is as high as the design temperature.

18.8.1 Decision on “Design Pressure @ Design Temperature” Pair

See Figure 18.26, which shows pressure and temperature changes in a piece of equipment during its life time. Pressure and temperature go up and down.

“Bubble 1” shows the absolute highest temperature. “Bubble 2” shows absolute highest pressure.

From a pure theoretical view point the pair of “design pressure @ design temperature” should be selected as the pair inside of “bubble 3.” This is because it represents the highest pressure at the highest temperature. However, we generally and negligently report “bubble 4” as the pair of “design pressure @ design temperature.” “Bubble 4” is obviously not a coincident pair.

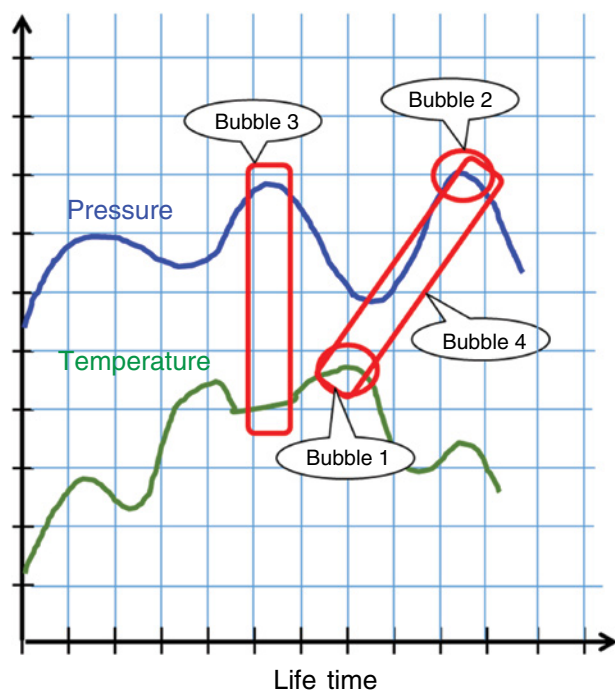


Figure 18.26 Pressure–temperature pair fluctuations in a piece of equipment.

In the next two sections we discuss decision process for selecting design pressure and design temperature.

18.8.1.1 Deciding on “Design Pressure”

If you remember, in the level system of pressure mentioned in Chapter 5 the design pressure is a level of pressure higher than the HHP (high-high pressure) and the HP (high pressure) and NP (normal pressure). For the selection of design pressure we can go high, very much higher than the HHP as much as we want, but this increases the cost of process elements. Therefore we need to bring down the design pressure to a level that is inexpensive while safe.

The way we define the “process design” is firstly define it through a minimalistic approach and the increase is to go higher than (or equal to) the HHP.

What is the minimum safe level of pressure that could be selected as design pressure?

As it is attempted to keep the pressure on and controlled at the normal pressure (NP) one may say the design pressure could be placed at the normal pressure! However the controllers that try to bring the pressure to the normal pressure are not perfect. Control loops have an “overshoot.”

In nutshell, “overshoot” is a magnitude of deviation from the pre-determined set point of a controller when it tries to keep the parameter at the set point.

As the control technologists generally adjust an overshoot of 10% in process plants, the design pressure could be selected as 10% higher than normal pressure as minimum.

After preliminary selection of design pressure as “NP + 10%” we should check to make sure the design pressure is higher than the HHP and, if it is needed, increase the design pressure to make sure the rest of the pressure levels, HHP and HP, are laid down somewhere in a band between NP and design pressure.

18.8.1.2 Deciding on “Design Temperature”

As was mentioned, the design temperature in the pair of “design pressure @ design temperature” should be a coincident value. However this temperature is generally decided independently of the design pressure. This is the meaning of selecting “bubble 4” in Figure 18.26. What we did basically is selecting the highest absolute pressure and selecting the highest absolute temperature and tied them “nominally” into a “pair.” We chose this selection and not the more accurate pair of “bubble 3” just because it is easier to do that. It is very difficult to estimate the maximum upset temperature at the upset pressure.

The selected pair may call for stronger, more expensive equipment but if the additional cost is acceptable, this easier methodology can be used.

However if non-coincident “design pressure @ design temperature” is selected one check should be done to make sure nothing is missed.

This check is: “the pressure that an item can tolerate at the highest absolute temperature should be checked too.” This check basically means making sure that the pressure corresponding to the non-coincident design temperature is not higher than the selected design pressure.

The design temperature of an item could be decided based on a specific margin on the HHT (high-high temperature) of the item. The margin could be any number from 5 to 30 °C and is instructed by the company guidelines.

18.8.2 Sources of Rebel Pressures

Design pressure is decided based on rebel pressure. However there are some cases that rebel pressure cannot be used for the purpose of design pressure specification. These are the cases where there is no specific maximum sustainable rebel pressure. For example if the discharge side of a positive displacement valve is closed off, the pressure will rebel and increase. However as this pressure doesn't eventually stay at a specific value, it cannot be used as the design pressure. Such cases can only be handled by placing a pressure safety device, as stated in Chapter 12.

In Table 18.11 some reasons for rebelling pressures, either on the high pressure side or low pressure side, are listed.

Here we explain one important rebel high pressure and one rebel low pressure scenario.

The rebel high pressure can be decided to be the “dead head pressure or shut-off pressure.” It is very common to see the “design pressure” of items on the discharge side of a dynamic fluid mover if it is decided based on the “dead head differential pressure” of the fluid mover.

Table 18.11 Specific rebel pressure scenarios.

Rebel high pressure	Rebel low pressure
<ul style="list-style-type: none"> Centrifugal fluid mover discharge closing off: dead head pressure Fail open of control valve or regulator where it is connected to high pressure reservoir Runaway reaction where the products of side-reactions are gaseous Vaporization (because of abnormal heat) 	<ul style="list-style-type: none"> Fail open of control valve or regulator where it is connected to vacuum reservoir Runaway reaction where the product of side-reactions are liquid while the raw material are gaseous Vapor condensation (because of abnormal cooling)

The rebel low pressure can be decided to be “full vacuum” for the equipment that may need “steaming out” during its life cycle in the plant. If a piece of equipment deals with oily material it may need steaming out for cleaning purposes. It has been observed before that a vessel was cleaned by steaming-out, and a quick but harsh rain caused quick condensation in the vessel, which led to a vacuum inside of the vessel and the vessel crumpled like a piece of paper. For such cases the design vacuum of the vessel should be specified as “full vacuum.”

18.8.3 Sources of Rebel Temperatures

Listing the scenarios for rebel temperatures are easier where there is something that change the temperature of the process fluid.

For example in an exothermic reactor, the HHT could be decided based on the maximum temperature attained when the cooling water to the reactor jacket is – for whatever reason – stopped.

The other example is the HHT for a piece of equipment downstream of two heat exchanger in series may be decided when one heat exchanger (possible the one with larger duty) fails to do its functionality.

However there are some other cases that rebel temperatures exists because of other reasons.

In Table 18.12, some reasons for rebelling temperatures, either on the high temperature side or low temperature side, are listed.

18.8.4 Design Pressure and Design Temperature of Single Process Elements

Below the design pressures of several items are discussed.

- Tank: the tank design pressures are requested by process engineers and are provided by the mechanical engineers of the fabricator. The requested design pressures, however, should be less than the maximum allowable design pressure dictated by the associated

Table 18.12 Specific rebel temperature scenarios.

Rebel high temperature	Rebel low temperature
<ul style="list-style-type: none"> Runaway exothermic reaction Fail open of control valve or regulator where it is connected to high temperature fluid 	<ul style="list-style-type: none"> Runaway endothermic reaction Fail open of control valve or regulator where it is connected to high temperature fluid Quick vaporization (because of pressure drop) Quick pressure drop of a liquefied gas because of Joule–Thomson effect (in some cases temperature increases)

tank design standard. For example the design pressure of a tank based on API 650 cannot be higher than a specific value.

- Vessel: the pair of “design pressure, design temperature” for vessels are provided by the fabricators and generally stated in the vessel call-out on the top or bottom of the P&ID sheet.

The important point is that the design pressure value of vessels refers to the maximum pressure that can be tolerated for the *top* of the vessel.

- Other equipment: for equipment the pair of “design pressure, design temperature” is generally stated in the equipment call-out at the top or bottom of the P&ID sheet. This pair is provided by the vendor. The important point is that the pair of “design pressure, design temperature” is stated for the piece of equipment in static conditions. This means that if there is a rotating shaft in the equipment, the rotating shaft and the associated items may blow out and dislocate pressures even below design pressures.

- Pipe spools: the design pressure and design temperature pipes are not as clear as other items, which means more caution needs to be used when dealing with pipes.

Pipe spools include pipes and flanges. The weakness points of pipe spools are the flanges. Because of that when the design pressure of a pipe spool is mentioned, the design pressure of the connected flange is also mentioned.

The pair of “design pressure, design temperature” of flanges are mentioned in the form of “pressure rating,” or “rating.” For example the “design pressure, design temperature” of a pipe spool could be #300. This means the weakest points of this spool, which are its flanges, have a rating of #300. By knowing the material of the flange, a related reference table shows what would be the design pressure of the spool at its design temperature. For example if the material of the spool flanges are group 1.1 of ASME B31.1 and the design temperature of the spool is 80 °C, the design pressure can be read from the table and would be 2000 KPag. It is important to know that this design pressure is for the spool or the flanges, the pipe by itself may have a design pressure seven or eight times higher than the spool design pressure.

Defined classes are 25, 125, 250, and 800 for cast iron; 150 and 300 for malleable iron; and 150, 300, 400, 600, 900, 1500, and 2500 for steel.

The rating of flanges can be accepted as a meaningless number that only ranks the robustness of flanges. However, this rating was known as the “pressure rating” in older days. The “pressure rating” could be roughly assumed as the design pressure of a flange (in psi) at a design temperature of 750–850 °C. The importance of the temperatures of 750–850 °C is that it is the temperature at which the majority of steels start to yield.

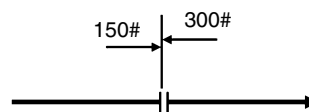


Figure 18.27 Pipe rating border.

If the piping material spec of a project is not available and we are looking for accurate value, a rule of thumb can be used to get the design pressure of conventional carbon steel (in KPa): in ambient temperature (20–40 °C) it is about 16.5 times the flange class.

In P&IDs the pipe spool rating can be recognized from the pipe tag. The pipe spec appears in the pipe tag as an acronym, which represent the “piping spec”. However, where a pipe spool rating is changed throughout a long pipe route, the rating is mentioned on each side of the rating change point on the piping. This writing helps to identify the exact location of the rating change. One example is shown in Figure 18.27.

The rating of pipe spools is primarily determined by the process. However, there would be a minimum rating for a piping spool to sustain its integrity. Generally the minimum acceptable rating is #300 for small bore spools (say less than 3”). The same logic applies to control valves. The pressure rating of small control valves (say less than 3” or 4”) is a minimum of 300#.

18.8.5 Design Pressure of Connected Items

The question arises during P&ID development when two elements (or more) are connected to each other and their design pressure before connecting them together was different. What should be the design pressure of connected items? Can they be different or they should be equal?

Figure 18.28 shows the pressure change in a typical plant along the “string.”

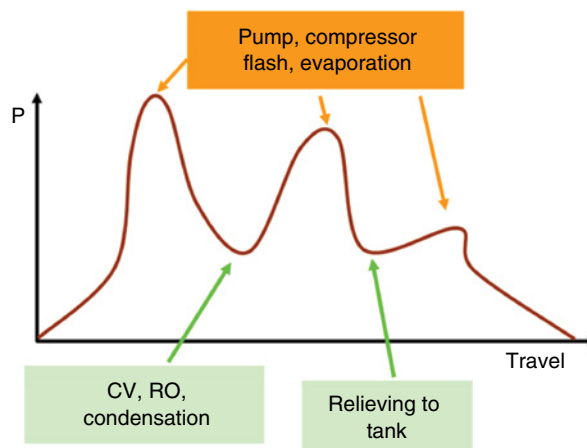


Figure 18.28 Pressure change in a string of units.

In a tunnel vision, a process engineer may select the design pressure of each unit based on the operating pressure of the unit. However this narrow sighted decision may lead to big hazards.

It is a good idea that the design pressure of connected items is checked in a broader sense by the process engineer.

This concept is explained below for two common cases of “connected equipment to each other” and “connected equipment to instrument.”

18.8.5.1 Design Pressure of Connected Equipment–Equipment

What we need to do when we connect two (or more) pieces of equipment together while their operating and design pressures are different than each other?

There are basically two solutions available: one is limiting and allocating the design pressure to each unit, and the second one is equalizing the design pressure of all the connected units to the highest value.

Figure 18.29 shows a string of units with different operating and design pressures.

However this schematic is a simplistic version of reality. In reality, during PFD and P&ID development this string is provided with a pressure decreasing system, pressure regulator or pressure decreasing control valve, as shown on Figure 18.30.

The point here is to make sure the high pressure doesn't sweep away from upstream to downstream; in a low pressure system, a safety valve should be placed downstream of each pressure reducing system (Figure 18.31).

Placing a safety valve after a control valve is one common solution when connecting two pieces of equipment with two different operating and design pressures. The other solutions are placing two control valves in series (or two regulators) with the same set point, so that one of them is functioning as a control valve and the other one is monitoring the control valve. During operation, if one of the control valves fails, the other one automatically takes care of its duty. When both of them are operating, the second control valves doesn't really do anything to the flow though.

The other solution is placing a switching valve downstream of the pressure reducing control valve.

These solutions are shown in Figure 18.32.

If the designer is not willing to “allocate” and limit the design pressure to each unit through the above solutions

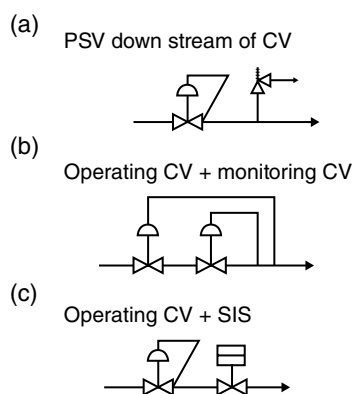


Figure 18.32 Different methods for limiting pressure to each unit.

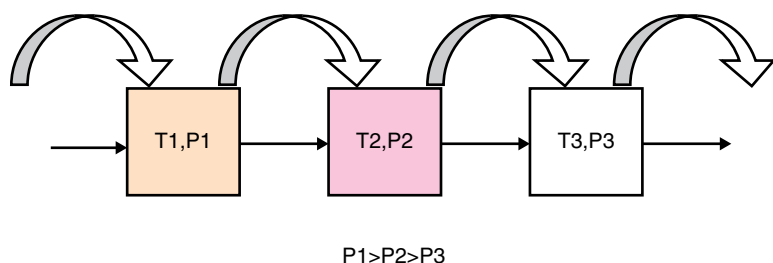


Figure 18.29 Design pressure of units in a string.

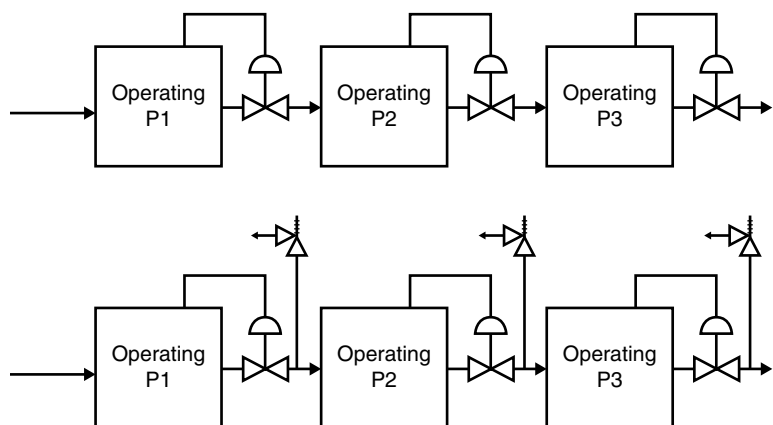


Figure 18.30 Units in a string with different design pressures.

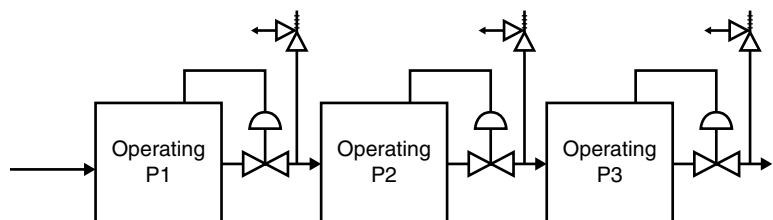


Figure 18.31 Limiting each pressure for each unit in a string by adding a pressure safety valve.

he may need to choose the inherent solution. In the inherent solution all the design pressures of connected items are equalized to the highest value. This is a very conservative approach that not all companies welcome because it may increase the cost of project to a high, unacceptable value. For example one company may say: “all the tanks connected to a VRU system through a vapor collection network should have the same design pressure (@ design temperature).”

Although this logic cannot be completely overruled, it is less likely to happen in a large scope and it is also very expensive to implement.

A more common example of where this situation happens a lot during P&ID development is what is shown in Figure 18.33.

What should we do when tying-in two pipes together with two different design pressures?

It is obvious that the operating pressure of the two separate pipes will both be changed to new values after the tying in. In an inherent solution the design pressure of the lower rating pipe should be increased to the higher rating (Figure 18.34).

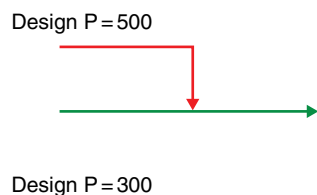


Figure 18.33 Tying together two pipes with different ratings.

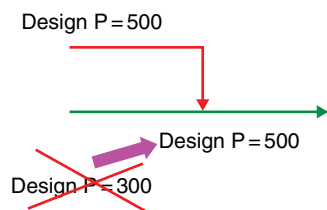


Figure 18.34 Equalization of pipe rating after tying them in together.

If this solution is not acceptable from an economical viewpoint the limiting pressure can be implemented. In this solution a pressure regulator together with a PSV can be placed on the lower rating pipe, right after the tying in point. A single or double check valve could be placed to prevent high pressure to “migrate” to the upstream of a lower rating pipe.

Placing a check valve for this purpose is very tricky and is not always acceptable. The reason is that a check valve may prevent reverse flow but not the reverse pressure! Then putting in a check valve doesn’t necessarily prevent high pressure from reversing to upstream of the tying point.

This lack of reverse pressure prevention is because no check valve can 100% prevent backflow. Generally speaking a conventional swing check valves passes flow in the reverse direction in about 10% of main flow.

18.8.5.2 Design Pressure of Connected Equipment–Sensor

For connected equipment–sensor, there are at least two available solutions: equalizing the design pressure to the highest value and do nothing!

The solution of limiting and allocating pressure is generally not available for this case.

Some companies prefer to put the design pressure of all instruments connected to – for example – a piece of pipes equal to the rating of the pipe. This could be an expensive approach and not all companies like it. It is not strange to see that the design pressure of connected instruments to a piece of pipe is lower than the rating of the pipe with no means of pressure limitation and with no concerns!

This “do nothing” approach could be taken by some companies based on the logic that: “losing the pipe and rupturing it is not affordable by us but we can afford to lose a small instrument if the pressure goes beyond the design pressure of the instrument.”

Therefore the answer to the question of: “should the design pressure of a flow sensor be the same as the design pressure of the pipe it is installed on it?” The short answer is: “not necessarily.”

19

General Procedures

19.1 Introduction

The purpose of this chapter is to show a general methodology for the development of P&IDs and a general methodology for checking P&IDs.

19.2 General Procedure for P&ID Development

The question is how to develop the P&ID of an item that is new to you.

Let's look at a process plant from a bird's eye view (Figure 19.1).

The main units in process engineering are: conversion units and separation units. The conversion units could be physical conversion units or chemical conversion units.

The other items, that we name general items, can be considered as "peripheral" items, and their duties satisfy the main conversion units.

We have already learned how to develop P&IDs for general items like pipes, pumps, compressors, heat exchangers, etc. However, it is not always the case that a design engineer (in the role of P&ID developer) should develop the P&ID for general/popular items (e.g. containers, fluid movers, heat exchangers, etc.). In those cases where he is faced with new items (less popular items like a liquid extraction tower, filter press, etc.) he should have the capability of developing the P&ID. It is not very easy to develop the P&ID when you are totally unfamiliar with the item, but it is not impossible.

The first step is to learn the function of the new piece of equipment and its principles of operation. Talking to the vendor and several users of the piece of equipment helps a lot in developing a good P&ID for a piece of equipment. Interviewing vendors is easy because they want to sell their equipment to you but finding good users for the equipment is not easy.

First of all users are generally hesitant to talk because the transferred information may be considered as

proprietary information and inhibited. The second issue is that getting unbiased information is very difficult. The third issue is every user's experience is gained in a specific service, specific weather, etc. and may not be considered as a "general" idea. In the end, it is the skillfulness of the P&ID development engineer to "extract" the pure facts from the interviews.

P&ID development is nothing but developing provisions to cover all four stages of the life cycle for every single piece of item on the plant.

These four stages are, again, normal operation, non-routine operation, maintenance/inspection, and the absence of the item from operation.

Here we develop this strategy in two sections, a piping and equipment section and an instrument, control, alarm, and SIS section.

19.2.1 P&ID Development: Piping and Equipment

Out of the four stages of each piece of equipment, the "normal operation stage" generally doesn't need much from a piping viewpoint. The majority of items are needed for the three other phases of operation.

Each of the items needed to be added to cover these four stages may need an additional control system.

For the piping and equipment section these sample questions could be asked:

- 1) The equipment may need partial recycling if the function of the equipment improves because of the recycling. Examples of such equipment are reactors with equilibrium reactions and the units where "probability" is a factor (like the floatation process in mining).
- 2) How low can the flow rate be for the equipment to work comfortably, and is there any expectation that the flow will go below that "minimum acceptable flow"? What happens if the flow goes below the "minimum acceptable flow" in the short and long term?

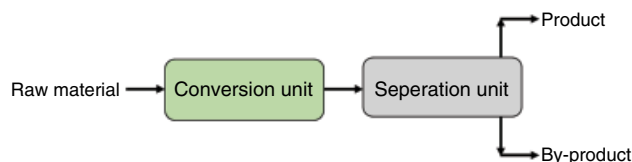


Figure 19.1 Process plant, a bird's eye view.

- 3) Think about potential weakness points of the equipment. Consider these words: thin, tight, non-metallic, multi-component part, expensive part, moving part, vibrating part, etc.
- 4) Is the equipment sensitive toward suspended solids and are there suspended solids in the incoming flow? If the answer is yes, a strainer may be needed upstream of the equipment.
- 5) Is the equipment sensitive toward surged flow and is the incoming flow likely to surge? If the answer is yes, a surge dampener may be needed upstream of the equipment.
- 6) What is the plan if an off-spec product is produced? Can it be recycled or should it be discarded?
- 7) What do you need to put in for easier inspection of the equipment? You need to put more facilitating things if more frequent inspection is needed or if missing an inspection has large consequences.
- 8) You most likely need to put an isolation system (including isolation valves) around your equipment unless you can not afford pulling off the operation of the equipment during the normal operation of the plant.
- 9) What do you want to do with incoming flow when the equipment is non-operational?

19.2.2 P&ID Development: Control and Instruments

After developing the piping and equipment portion, the next step is the instrumentation and control portion. However, after finishing this task you may need to check the piping and equipment portion again. P&ID development, like other design processes, is not a straightforward process.

Here again we need to cover the four stages of operation. However instrument and control are mainly needed for the first two stages: normal operation and non-routine operation of equipment.

For easy decision making a matrix similar to that shown in Figure 19.2 can be developed for each piece of equipment. The different process and non-process parameters are given in the first column and the different wrapping layers are placed in the first row. A check mark shows if it is decided to use each wrapping layer around each parameter.

Decision making for each steering loop was discussed in detail in Chapter 16.

ICSS action Parameter	Monitoring		Regulatory control	Interlock
	Field	Control		
Pressure				
Temperaure				
Flowrate				
Level				
Composition				

Figure 19.2 Steering component selection matrix.

However, here we provide a simple method as a preliminary step.

To decide about parameters for each steering loop, all the applicable parameters should be classified in five groups of “barely important parameters,” “mildly important parameters,” “very important parameters,” and “critical parameters.” Each importance level is connected to an I&C requirement: “nothing” for the first group, “monitor field” for the second group, “monitor control room” for the third group, “regulatory control loop” for the fourth group, and “safety interlock function” for the last group. Such correspondence is shown in Table 19.1.

There are some cases that a parameter is not definable for a piece of equipment. For example, for a vessel flow rate is not definable. In some other cases a parameter is not important. For example composition generally is not important for pumps.

Barely important parameters are chosen based on screening of the non-important parameters and mildly important parameters.

Mildly important parameters are the parameters that affect the operation of the unit but are not the main parameters of the unit.

Very important parameters are the parameters that are related to the main duty of the unit.

Critical parameters are the parameters for which their violation creates risk to personnel, assets and the environment. The increased risk could be through increasing the probability or consequences or both.



















Table 19.2 shows a parameter matrix for a typical pump in a hot water service.

In Table 19.2 different process parameters are examined against their criticality to come up with the required monitoring and control system. We generally don't care about the temperature in pumps but as this pump works with hot water then there is chance of having cavitation when the temperature is high. It is a good idea to put a

Table 19.1 Correspondence of importance level versus I&C requirement.

Not applicable parameter or no important parameter	Barely important parameter	Mildly important parameter	Very important parameter	Critical parameter
Nothing!	Monitor – field	Monitor – control room	Regulatory control loop	Safety interlock function

Table 19.2 An example of a parameter matrix for a pump in hot water service.

Water transfer pump 101-P-1287					
Importance	Not applicable parameter or no important parameter	Barely important parameter	Mildly important parameter	Very important parameter	Critical parameter
Required steering loop	Nothing!	Monitor – field	Monitor – control room	Regulatory control loop	Safety interlock function
Temperature			Yes, needs field indicator		
Pressure		Yes, needs pressure gauge on suction side	NA		Yes, needs SIF loop on suction side
Flow rate				Yes, needs Control loop (on discharge side)	
Level	Not applicable				
Composition	Not important				

pressure gauge to check the temperature in the field and also an interlock to stop the pump when there is not enough pressure at the suction of the pump. Where there is trip on a parameter, the parameter most likely can be “seen” in the control room too.

The flow rate and discharge pressure are the primary parameter fluid movers but for pumps the only primary parameter is the flow rate. Then a control loop is needed for the flow rate. Again here flow can be “seen” in the control room too.

Level is not definable for pumps.

Composition is not important for pumps.

Finally, two important points regarding the required monitoring and control system should be mentioned:

- 1) The parameters that have a control loop or SIF loop (very important parameters and critical parameters) should also be provided with control room monitoring.
- 2) In a noticeable number of cases if a parameter has control room monitoring, field monitoring is provided too.

Table 19.3 Quick and dirty decision-making on control architecture.

Affecting parameters: P_1, P_2
<ul style="list-style-type: none"> • P_1 is the most important (and only) parameter: single loop control • P_1 is an important parameter in the beginning, and then P_2 will become important: override control • P_1 is an important parameter in the beginning and then P_0 (a pre-specified number) will become important: limiting control • P_1 & P_2 are equally important, but P_1 works quicker than P_2: cascade control (P_1 & P_2 have some dependency on each other) • P_1 is the most important (only) parameter but it cannot be “managed” by one control valve: split-range control or parallel control

Here a quick decision making tool is provided for the control system.

- Guideline 1: Table 19.3 helps in choosing the best control architecture in a “quick and dirty” way.

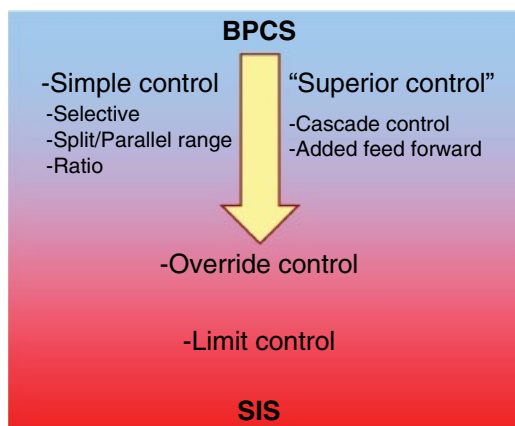


Figure 19.3 Moving from a BPCS toward a SIS.

- Guideline 2: Figure 19.3 shows the relative position of different control architectures in the framework of BPCS and SIS structures.

A simple control may need a single loop control, or, depending on the situation, selective control, split-range/parallel-range control, or ratio control.

If you are looking for “superior” control (tight control) you may choose feedback + feedforward control, or cascade control, or even feedback + feedforward + cascade control.

When the control action starts to deviate from regulatory control (or BPCS) and to go toward a more trip-type control (or SIS type), override control, and then limit control, can be implemented (Figure 19.3).

From a very simplistic point of view, it could be said that: “since the purpose of a unit operation/process unit is to convert one material into another – physically or chemically – the only required control loops are ‘composition control loops.’” This viewpoint is generally incorrect. In the majority of cases, we control unit operations and unit processes, not through composition loops, but through other loops. We mentioned before that for various reasons, we prefer not to use composition control loops unless we really have to.

Instead, we try to find some “underlying parameters” (among temperature, pressure, flow rate, and level) that are known to “direct” the composition, and then we control those simple parameters.

The above statement is the golden rule in controlling unit operations and unit processes. One imaginary composition loop may be broken into a few T/F/P/L loops. This could raise the issue of the loops interfering with each other.

Beware the majority of units, either conversion units or separation units, that have a portion to store or hold fluids. This means the control of each unit most likely includes some aspects of vessel or tank control.

Actually, container control is one common control scheme in the majority of unit operation and unit process controls.

The below are some general rules that help:

- 1) The equipment that we buy for a plant is not “custom built equipment” that we can then expect to operate exactly to our operating needs. Even for the case of custom made items, we usually expect equipment to operate in a pre-determined “window” of operation. The result is that almost all equipment in the plant should be “tamed” through a control system. Provide the required control (BPCS) to bring about the duty of the item. The item will be bought to do a task but usually the item will output a range of parameters. BPCS control will force the equipment to function in the required “window.” Examples of tasks are flow rate and head for pumps, and heat duty for heat exchangers.
- 2) Check the required temperature, pressure for the item (inlet, outlet)
- 3) Check the required flow rate for the item. What is the minimum flow rate that can be handled without impact on process and what is the minimum flow before there is harm to the equipment?
- 4) Check the required composition for the item and care that should be taken. For example, a positive displacement pump is very prone to plugging if liquid has large suspended solids. In this case a strainer should be placed.
- 5) What are the required utilities and their temperature and pressure?
- 6) What are the weak points of the item and the care that should be taken in designing a proper SIS for the item?
- 7) Which parameters of the item need to be monitored by a rounding operator? (Think about those five parameters: temperature, pressure, level, flow rate, and composition.)
- 8) Which portions of the item need inspection and/or monitoring.
- 9) Is any history of item failures (frequency and time for maintenance, etc.) available? It can affect stages three and four of the item. For this step you may need to interview other users.
- 10) What if we lose the item? How can we minimize its impact on the rest of plant? Can we have a similar system as spare? (If the item is expensive other options should be considered.)

Now we are going to apply our learning to develop a P&ID for two pieces of non-general equipment:

- Example 1: gravity separator control (Figure 19.4)
- Example 2: flash drum control (Figure 19.5)

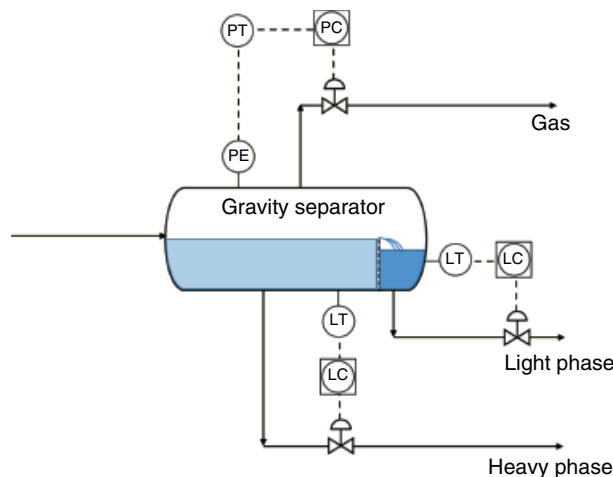


Figure 19.4 Developing a control system for a two phase liquid–liquid separator.

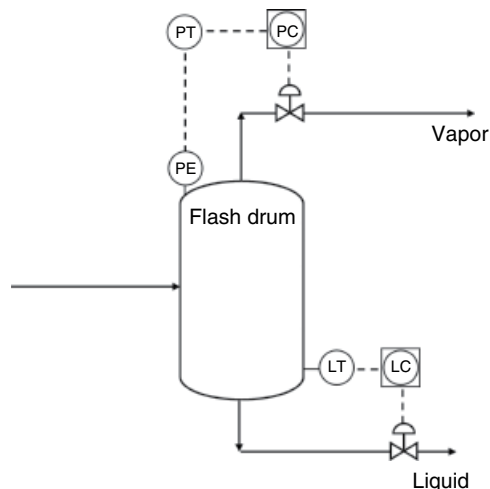


Figure 19.5 Developing a control system for a flash drum.

There are at least four levels of P&ID checking. They could be classified as below from the lightest to deepest check:

- 1) Format check
- 2) Demonstration rules check
- 3) Technical check
- 4) Design check.

In the format check things are checked that have no specific rules for them in the guidelines of the company (client or engineering company). There are mainly cosmetic items and English correctness. The format check is actually a sort of “first impression check.”

In the demonstration check P&IDs are checked against the “rules” mentioned in the guidelines of the company (client or engineering company).

In the technical check the P&ID development aspects are checked. In this type of check we need to make sure the provided items around each piece cover all the phases of it during the life cycle of the plant. Life cycles (as mentioned in Chapter 5) include: normal operation, non-routine operation, inspection/maintenance, provision for lack of items.

However, it is not easy to do that. Doing a mini-HAZOP in your mind for each item helps a lot in systematically doing a technical check. A HAZOP or mini-HAZOP, however, doesn’t cover the maintenance phase of process plants.

It is important to remember not only the main process P&IDs but other P&IDs like HVAC P&IDs, sampling system P&IDs, etc. should be checked in the technical check.

In the design check we are actually making sure it is “compatible” with all mother documents and sister documents. This means the P&IDs should be checked that they follow the higher-level documents and also the same-level documents. The higher level documents of P&IDs are design calculation documents and PFDs. The same-level documents are data sheets.

19.3 P&ID Reviewing and Checking

There are situations that a set of P&IDs should be checked to find mistakes, if any. It could be checking by a senior engineer before officially issuing a P&ID set, or checking of the P&ID set by the client representative.

A P&ID checker will not necessarily have all the capabilities of a P&ID developer. However a good P&ID checker is the one who has enough quality to develop P&IDs.

In this section we try to explain the methods of P&ID checking. To start, we need to know the meaning and scope of P&ID checking.

19.3.1 Format Check

- P&ID cosmetic checking:
 - Make sure no symbol/phrases “touch” or cross each other
 - Make sure the flows go from left to right
 - Make sure there are no overcrowded areas in the drawing sheet
 - Make sure equipment is evenly spread
 - Make sure the symbol sizes roughly follow real equipment size
 - Make sure there is a minimum number of line (pipes or signals) changes in direction

- Make sure there is a minimum number of pipe and signal crossovers
- Make sure the pipe appurtenances (including valves) are on the horizontal portion of pipes as much as possible.
- P&ID English Checking:
 - Wording points and fonts are consistent
 - Make sure all the writing is written in capitals (if this is the rule)
 - Make sure the spelling of English words is correct
 - Be careful about plural “s” for multiple similar units
 - Make sure there are no identical names for equipment.

19.3.2 Demonstration Rules Check

- Non-technical boxes check
 - Make sure names are correct (e.g. project name, P&ID name, client name, signees name)
 - Make sure the revision number and issue date are correct
 - Make sure title block is accurate
 - Make sure disclaimer is accurate.
- Off-page connectors check
 - Make sure off-page connectors are horizontal as much as possible (in the utility drawing it is acceptable to see more vertical off-page connectors)
 - Make sure off-page connectors are near the edge of sheets.
- Symbol check
 - Make sure all symbols are correctly depicted based on the legend sheet (e.g. PCV versus PSV, centrifugal pumps versus PD pump, main pipe versus minor pipe)
 - Make sure all symbols are correctly depicted within the set of P&IDs
 - Make sure the type of dashed line are suitable for each application (for example the dashed line for electrical signals is different from the border of vendor-supplied packages)
 - When lines cross, vertical lines break for horizontal lines except that instrumentation lines break for process lines. (This rule varies by company.)
- Tags check
 - Make sure all items are tagged: packages, equipment, SP items, instruments (pay special attention to “long” items)
 - Make sure the equipment tags are inside of equipment, if they are large enough, or below them
 - Make sure each pipe has one tag (no less, no more) on each sheet
 - Make sure pipe and valve identifiers are on the horizontal portion of pipes
- Make sure vendor-supplied items carry item tags near their border and are legible.
- Call-out check
 - Make sure call-outs follow the project standard
 - Make sure call-outs are correctly located
 - Make sure the proper units are used for parameters in call-outs
 - Make sure vendor-supplied packages also carry call-outs.

19.3.3 Technical Check

- P&ID set completeness check
 - Make sure all drawings are available including auxiliary P&IDs.
- Hold and note check
 - Make sure all notes and holds are still valid
 - Make sure each note in the note area has a mating note in the body of the P&ID sheet; “stray” notes are unacceptable. This should be checked for all “specific notes”
 - Make sure all general notes are applicable
 - Make sure the phrase “NOTE X” in the body of the P&ID is as close as possible to the corresponding item.
- P&ID connectivity check
 - Make sure off-page connectors are hand-shaking through different sheets
 - Make sure off-page connectors carry a suitable explanation on the top of them about the nature of stream or signal
 - Make sure off-page connectors carry the suitable TO/FROM explanation below them
 - Make sure the correct P&ID tag is mentioned inside of off-page connectors
 - Make sure references to other P&IDs are correct, including vendor/licensor document numbers.
- Border checks
 - Make sure the borders for vendor-supplied systems are correctly depicted
 - Make sure no building border is missed (I/B, O/B)
 - Make sure no ground border is missed (U/G, A/G)
 - Make sure no battery limits border is missed (U/G, A/G)
 - Make sure no discipline responsibility border is missed.
- General Technical check for all items
 - Make sure each item on the body of the P&ID sheet has all its identifiers and they are matching
 - Check to make sure the call-outs are complete, accurate and up to date.
 - Make sure each piece of equipment has drains, vents, steam-out connections if they are needed
 - Check to make sure the isolation system follows the guidelines (e.g. double-block and bleed valves)

- Make sure the equipment has pressure and vacuum safety devices if they are needed (legally and technically)
- Check to make sure no users are missed and that they follow the real piping designed by pipe modelers in utility drawings
- Check to make sure all auxiliary drawings are correctly referenced to the main drawings.
- Technical check for pipes
 - Make sure every pipe has a pipe tag (size, commodity name)
 - Make sure pipes have insulation and a tracing symbol if needed. Be careful about very cold or very hot streams, especially outdoors.
 - Make sure pipes have “DO NOT POCKET,” “SLOPE,” “MIN LENGTH” and other design comments if they are needed
 - Make sure pipes have “NNF” (normally no flow) (or INTERMITTENT) if there is a chance of confusion
 - Make sure the existence of pipe arrows are shown on each turn of the line and in the correct direction
 - Make sure the sizes of connecting pipes are matching or a reducer/enlarger exists
 - Make sure the size of reducers and enlargers matches the sizes of both sides of the pipes
 - Make sure spectacle blinds are correctly shown normally open/closed and they are on the right side of the valve (if a valve exist)
 - Make sure pipe specs are all shown and correct and spec breaks are shown correctly
 - Check all “T” connections to make sure pipes connect to each other according to the project’s branch table (e.g. do I need to show an expander before the tee?)
 - Make sure different cleaning connection/valves including steaming-outs, chemical cleaning, or flushing are provided if they are needed.
 - Check to make sure tie point tags are provided at pipe interfaces.
- Technical check for valves
 - Make sure the shown valves are available for the selected pipe specification
 - Check to make sure every valve has proper symbol (gate, globe, etc.) based on the legend sheet
 - Check to make sure every valve has the proper size and tag
 - Check to make sure every valve has proper connections (e.g. flanges) on valve inlets and outlets
 - Make sure manual blocking valves have a regular position (NC or NO) if there is a chance of confusion
 - Make sure manual blocking valves have car sealed open or closed, lock open or closed if they need to have
 - Check for additional valve types specifying abbreviations like FP (full port)
 - Check to make sure all automatic valves carry failure positions like FO, FC, FL
- Check the control valves to make sure they have a suitable arrangement for maintenance (e.g. block/bypass valves, reducers)
- Check to make sure the symbol of the control valve was changed from a general symbol to specific symbol for the selected type of control valve
- Check to make sure safety valves have the required information like name of the governing case, set pressure, orifice designation, and inlet/outlet sizes
- Make sure the safety valve inlet/outlet sizes match the connected pipes or suitable size reducers are shown
- Make sure a spec break exists on the outlet side of safety devices
- Check to make sure the inlet and outlet of the valves are full port (FP) and car sealed open (CSO).
- Technical check for fluid movers
 - Check to make sure temporary strainers or permanent strainers exist, if they are needed
 - Check to make sure drains and vents are available for the casing
 - Check to make sure electric motors are shown correctly with the associated controls
 - If the fluid mover is a steam turbine make sure it is shown correctly and steam is provided for it
 - If the fluid mover is air-driven make sure it is shown correctly and air is provided for it
 - Check to make sure the seal flush plans are shown correctly, if they are needed
 - Check to make sure the auxiliary drawings are correctly referenced, if they exist.
- Technical check for vessels and tanks
 - Make sure all the internals are shown completely and correctly
 - Make sure all the levels are shown correctly: HHLL, HLL, NLL, ...
 - Check to have drain/truck out nozzles
 - Check to have overflow nozzles
 - Check to have manways on shells and roofs
 - Check to have a nozzle number for all nozzles
 - Check to make sure the container has a heater, if it is needed
 - Check to make sure the container has agitators, if they are needed
 - Make sure the skirt height to grade is stated if it is necessary (for vessels)
 - Check to make sure a breathing valve and VRU connection are available for tanks.
- Technical check for heat exchangers
 - Check whether the heat exchanger needs isolation valves or not
 - Check whether the heat exchanger needs chemical washing valves
 - Make sure condensers/reboilers have elevation.

- Technical check for fired heaters and boilers
 - Check to make sure the fired heater/boiler call-out is complete, accurate and up to date
 - Check to make sure there are adequate devices for the rounding operator to check the flame shape.
- Technical check for instruments and control system
 - Check to make sure instruments are of the right type
 - Check to make sure instrumentation signal lines are of the correct line type

19.3.4 Design Check

The design check means checking to make sure the P&ID follows the other related documents. These “related documents” are generally interpreted as “tier 1” related documents. This means only the documents that are directly related to P&IDs are checked.

If the concept is stretched to tier 2 documents checking the sizing calculations and connectivity with heat and material balance tables could also be included. This is generally beyond the scope of P&ID checking.

The design check is in two levels, document connectivity check and cursory sizing check.

- Document connectivity check
 - Make sure the P&ID follows the PFD
 - Make sure the control system in the PDF is applied in the P&ID
 - Make sure mark ups on the previous P&IDs are implemented correctly (make sure no new set of mistakes is generated by the drafter because of non-set-up drawing software)
 - Check the drafting process did not introduce any random new errors or mistakes. (Sometimes the CAD software setting is not perfect and a new set) of issues are generated after fixing some other issues by the CAD drafter.
 - Make sure equipment names match other documents like the PFD, equipment list, datasheet, etc.

- Making sure information on the P&ID matches the equipment data sheet.
- Cursory sizing check
 - Use rules of thumb to check some sizing
 - Making sure the sizes of connected items (e.g. pipe to vessel) “look” correct.

19.4 Methods of P&ID Reviewing and Checking

A P&ID checker/reviewer can decide (or be instructed) to review a P&ID in any of, or a combination of, two methods: the systematic approach and the scanning approach.

In either case, it is a good idea to not start and finish the review in one day and/or by one person. To make sure all the problems of the P&IDs are captured we need some sort of “cold eye” every several hours. Otherwise the eyes of the reviewer get used to the elements of P&IDs and may not catch mistakes. Sleeping on a partially checked P&ID helps you to capture more mistakes the next morning.

If the deadline doesn’t allow, you may ask someone else to check them, as a cold eye reviewer, at the end.

19.4.1 Systematic Approach

In this method the checker checks each sheet of P&ID based on the check list he has in hand. This method works for everyone, including less experienced people, but it is time consuming. In this method, for each piece of equipment, pipe, or even instrument a check list is developed. One good source is API14C (Figure 19.6).

19.4.2 Scanning Approach

This could be a quicker method but for trained, good eyes. This is the method of choice for more experienced people who are not directly responsible for checking P&IDs but have to approve them.

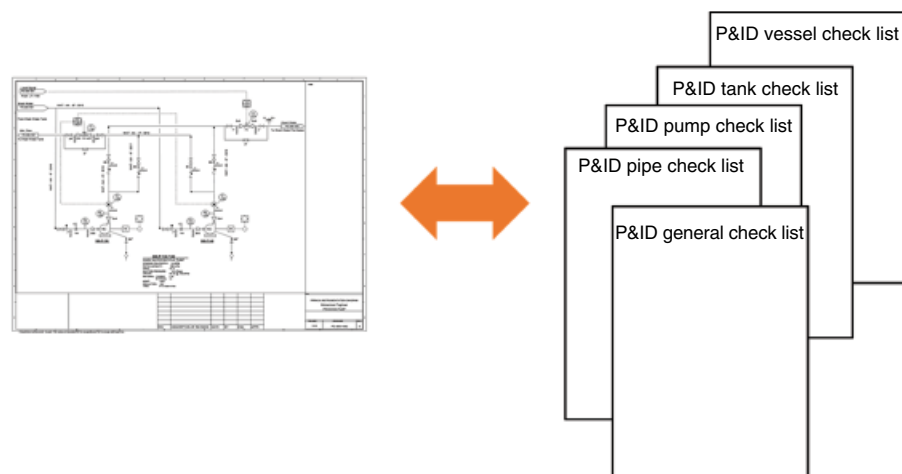


Figure 19.6 Example of check list using API14C.

In this method the brain and eyes should work closely together to do a type of “finding the hidden object” game!

In this method the checker scans the P&ID to find missing items and text, or illegibility.

19.5 Required Quality of P&IDs at Each Stage of Development

Now the question is what should be the quality of P&IDs at each step of a design project. For example do all the drain valves need to be shown even in the IFA revision of

a P&ID set? The answer to the above question is clearly no. To expand more on the answer, it should be noted that it is expected that more details are depicted on the P&ID when we are approaching the end of project. At the beginning of the project, for example on the IFR version of a P&ID, only large items are shown and no detail can be found. On the last revision of a P&ID, the IFC version, all the details should be depicted.

Each company has its own “standard” for quality of P&IDs in each stage of development. However, Table 19.4 can be used as a guideline where there is no standard available.

Table 19.4 Quality of P&IDs at each step of a design project.

		IFR	IFA	IFD	IFC
1.00	Equipment				
1.01	Positioning (necessity, existence)	Majority of them	All		Complete
1.02	Type	Majority of them	All		Complete
1.03	Tag	Majority of them	All		Complete
1.04	Call-out: capacity	Not available except for long lead items	All	Fine-tuned with vendor data	Complete
1.05	Call-out: other numerical specifications	X	Some of them	Majority of them	Complete
1.06	Required number of them and sparing policy	Majority of them	All		Complete
1.07	Materials of construction	Majority of them	All	Fine-tuned with vendor data	Complete
1.08	Diver – type		Majority of them	All	Complete (after fine-tuning)
1.09	Diver – power		Estimation	Majority of them	Complete
1.10	Critical elevations		Majority of them	All	Complete (after fine-tuning)
1.11	Utility positioning for equipment (requirement of utilities)		Some of them	Majority of them	Complete
1.12	Utility branch sizing for equipment		Some of them	Majority of them	Complete
1.13	Equipment isolation arrangement			Majority of them	Complete
2.00	Packaged units				
2.01	Positioning (necessity, existence)	Majority of them	All	Fine-tuned with vendor data	Complete
2.02	Type of components		Estimation	Fine-tuned with vendor data	Complete
2.03	Tag	Majority of them	All	Fine-tuned with vendor data	Complete
2.04	Call-out: capacity		All	Fine-tuned with vendor data	Complete
2.05	Call-out: other numerical specifications			Some of them	Complete
2.06	Required number of them and sparing policy	Majority of them	All	Fine-tuned with vendor data	Complete

(Continued)

Table 19.4 (Continued)

		IFR	IFA	IFD	IFC
	2.07	Materials of construction		Some of them	Complete
	2.08	Diver – type		All	Complete (after fine-tuning)
	2.09	Diver – power	Estimation	Majority of them	Complete
	2.10	Critical elevations		Some of them	Complete
	2.11	Utility positioning for equipment (requirement of utilities)	Some of them	Majority of them	Complete
	2.12	Utility branch sizing for the package	Some of them	Majority of them	Complete
	2.13	Package isolation arrangement		Majority of them	Complete
3.00		Pipes			
	3.01	Positioning (necessity, existence) – route	Some of them	Majority of them	All
	3.02	Size	All of 10" pipes and larger	All of 4" pipes and larger	All
	3.03	Tag	Few of them	Some of them	Majority of them
	3.04	Tie-in designations		Some of them	Majority of them
	3.05	Sloped (direction, value)		Slope direction	Slope value
4.00		Fittings			
	4.01	(Process) flanges, end fittings		Some of them	Majority of them
	4.02	Reducer (enlarger) – necessity, type, size		Some of them	Majority of them
	4.03	Tee	Few of them	Some of them	Majority of them
	4.04	Blinds			Majority of them
5.00		Manual valves			
	5.01	Positioning (necessity, existence) and type	Few of them	Some of them	Majority of them
	5.02	Size (the ones that are not the same pipe size)		Some of them	Majority of them
	5.03	Tag			Some of them
	5.04	Equipment block valves and bypass valves		Some of them	Majority of them
	5.05	Equipment drains, vents – positioning, size			Some of them
	5.06	Low point drains and high point vents			
6.00		Piping specialty items			
	6.01	Positioning (necessity, existence), type, size	Few of them	Some of them	Majority of them
	6.02	Tag			Majority of them
7.00		Instruments			
	7.01	Sensors – monitoring (out of loops)			Majority of them
	7.02	Sensors – part of loops	Some of them	Majority of them	All
	7.03	Control loops structure	Preliminary	Progress	Complete
	7.04	Safety instrument functions structure		Preliminary	Progress

Table 19.4 (Continued)

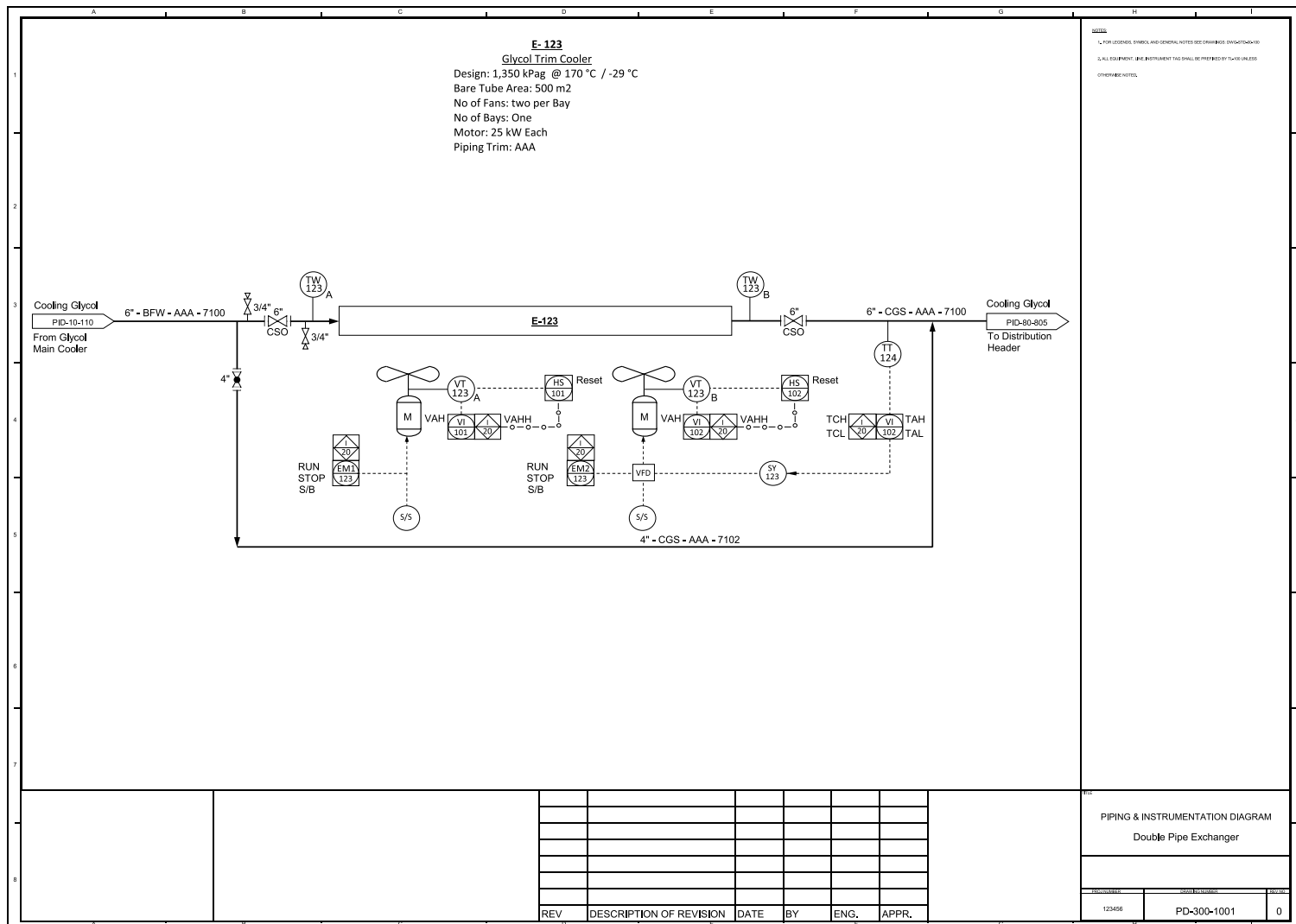
		IFR	IFA	IFD	IFC
8.00	Automatic valve (control and switching valves)				
8.01	Positioning (necessity, existence)	Few of them	Some of them	Majority of them	Complete
8.02	Type			Majority of them	Complete
8.03	Size		All of 4" and larger	All	Complete (after fine-tuning)
8.04	Tag		Majority of them	All	Complete (after fine-tuning)
9.00	Relief devices				
9.01	Positioning (necessity)	Majority of them	All	Fine-tuned with vendor data	Complete
9.02	Governing scenario		Majority of them	All	Complete (after fine-tuning)
9.03	Type		Some of them	Majority of them	Complete
9.04	Size		Some of them	Majority of them	Complete
9.05	Tagging			Majority of them	Complete
9.06	Relief device inlet/outlet size		Some of them	Majority of them	Complete
10.00	Insulation/tracing				
10.01	Positioning (necessity)		Some of them	Majority of them	Complete
10.02	Specification (type, thickness)			Majority of them	Complete
11.00	Utility networks				
11.01	Utility distribution and collection network		Preliminary	Complete	Complete
11.02	Flare header collection network		Preliminary	Complete	Complete
11.03	Drain/vent collection system			Complete	Complete
12.00	Safety shower/eye washer			Preliminary	Complete
13.00	Utility station			Preliminary	Complete
14.00	Fire and gas detection system			Preliminary	Complete
15.00	Firefighting system and deluge system			Preliminary	Complete
16.00	HVAC system (auxiliary P&ID)			Preliminary	Complete
17.00	Pump seal flush system (auxiliary P&ID)			Preliminary	Complete
18.00	Lube oil system (auxiliary P&ID)			Preliminary	Complete
19.00	Sampling system (auxiliary P&ID)			Preliminary	Complete
20.00	Sensor installation detail (auxiliary P&ID)			Preliminary	Complete
21.00	Automatic valve detail (auxiliary P&ID)			Preliminary	Complete

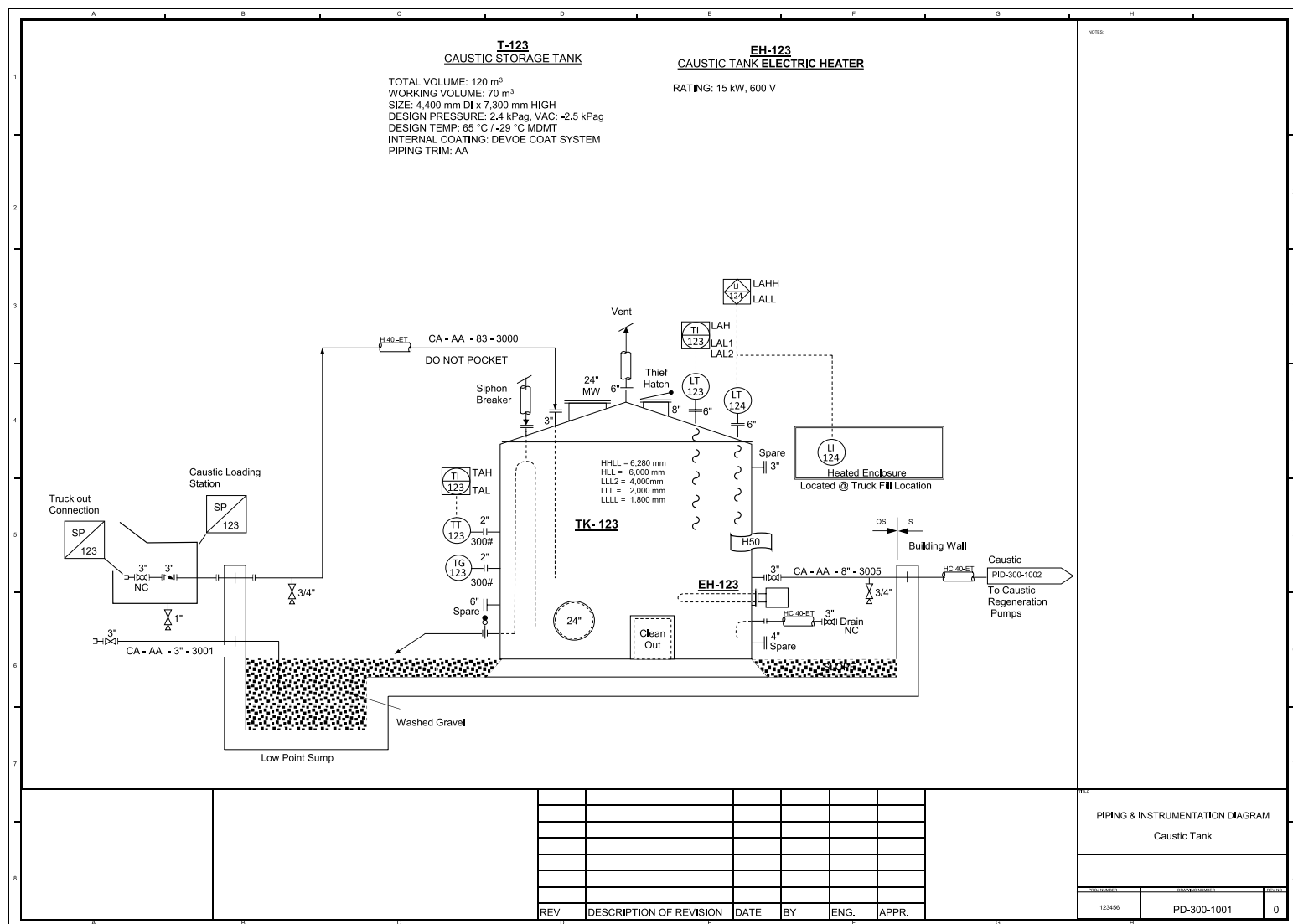
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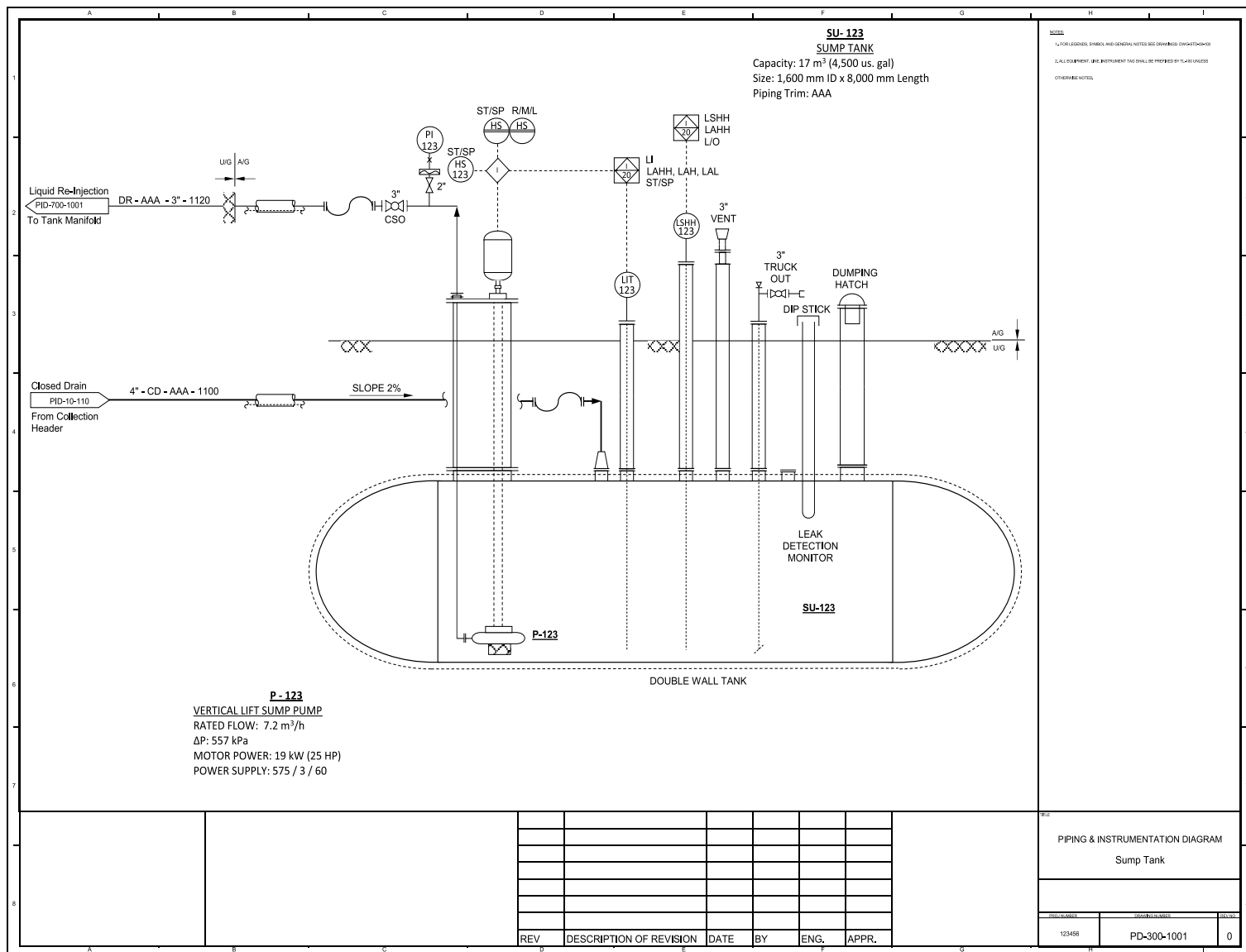
Examples



Air Cooler







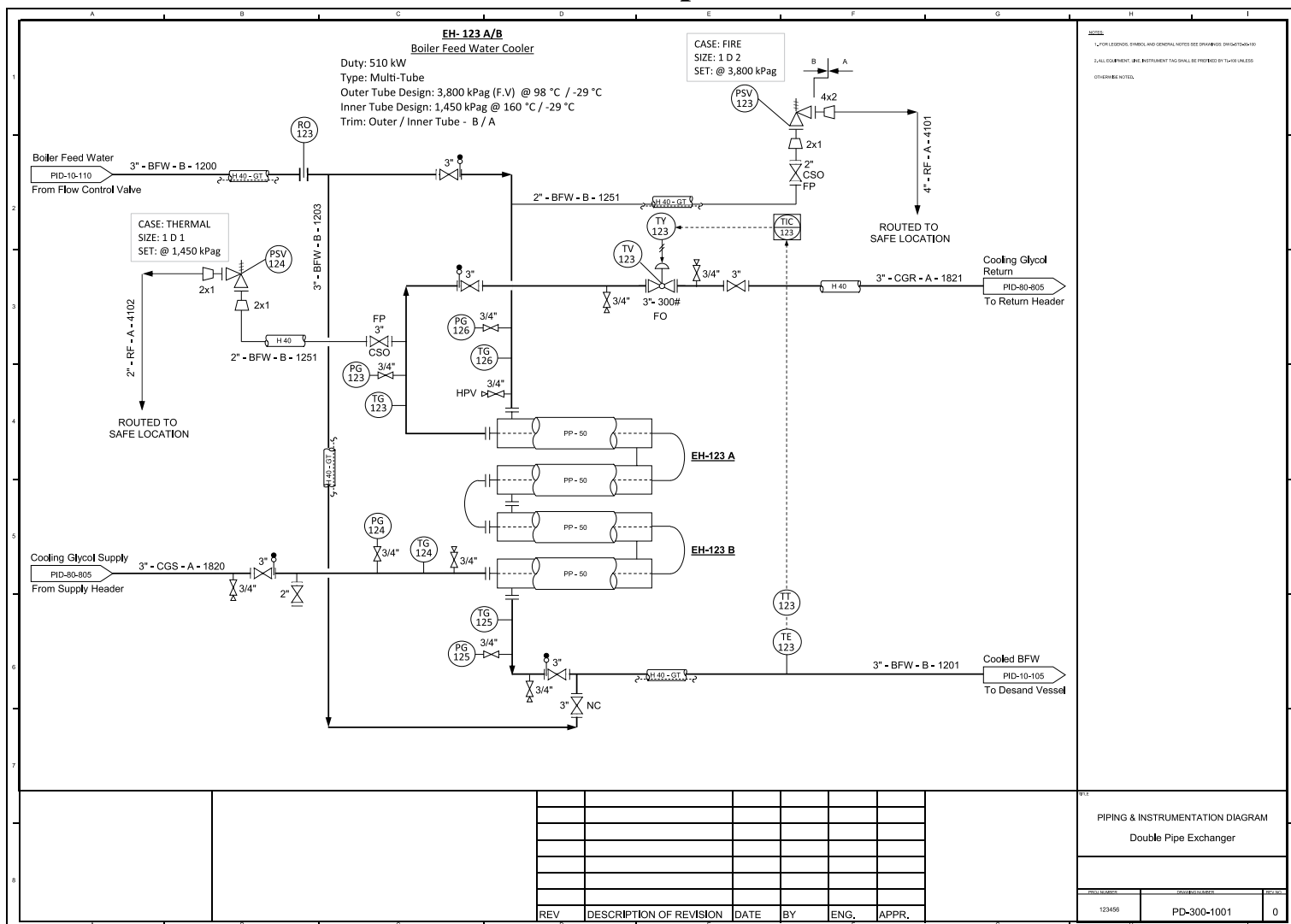
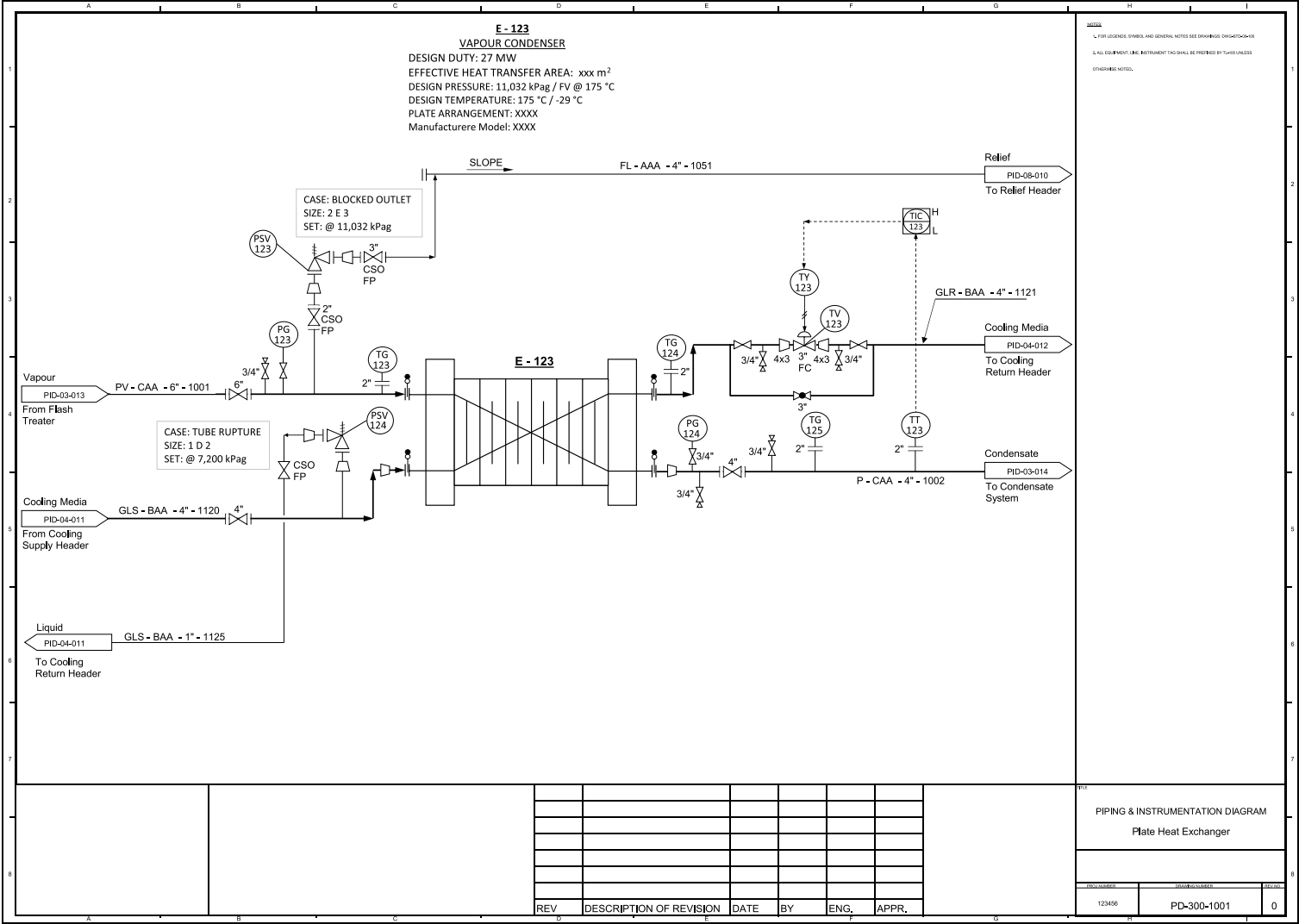
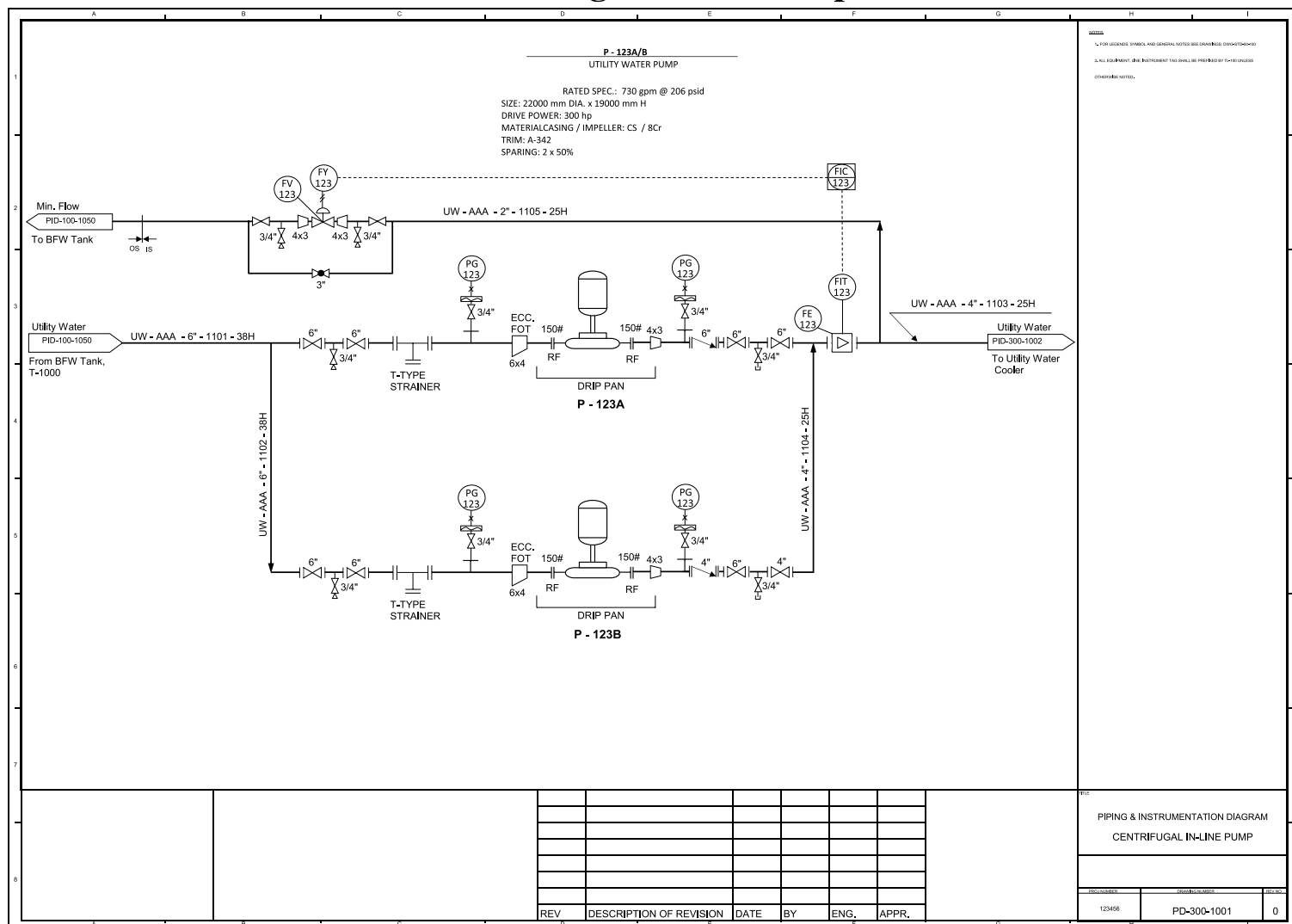
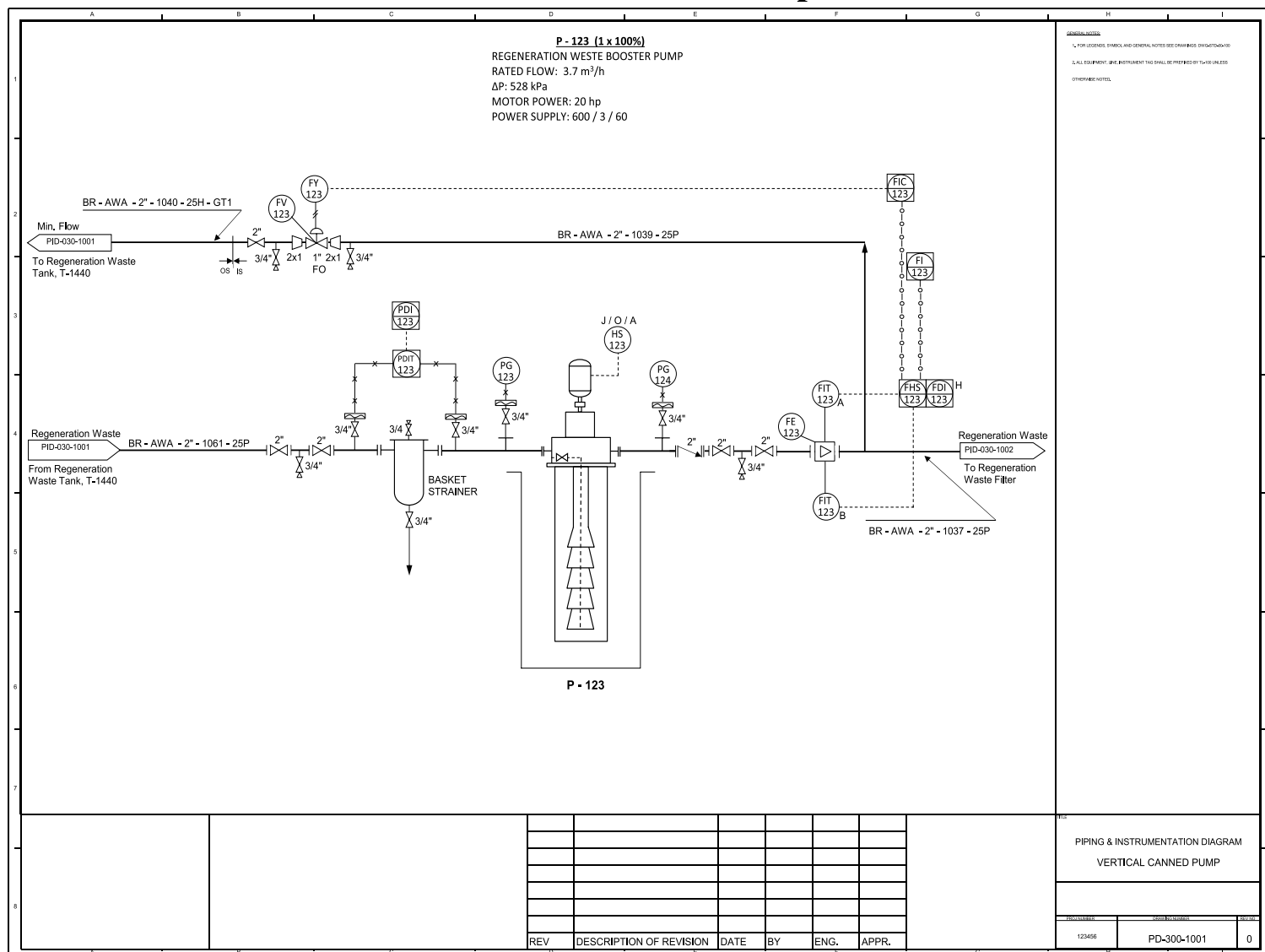


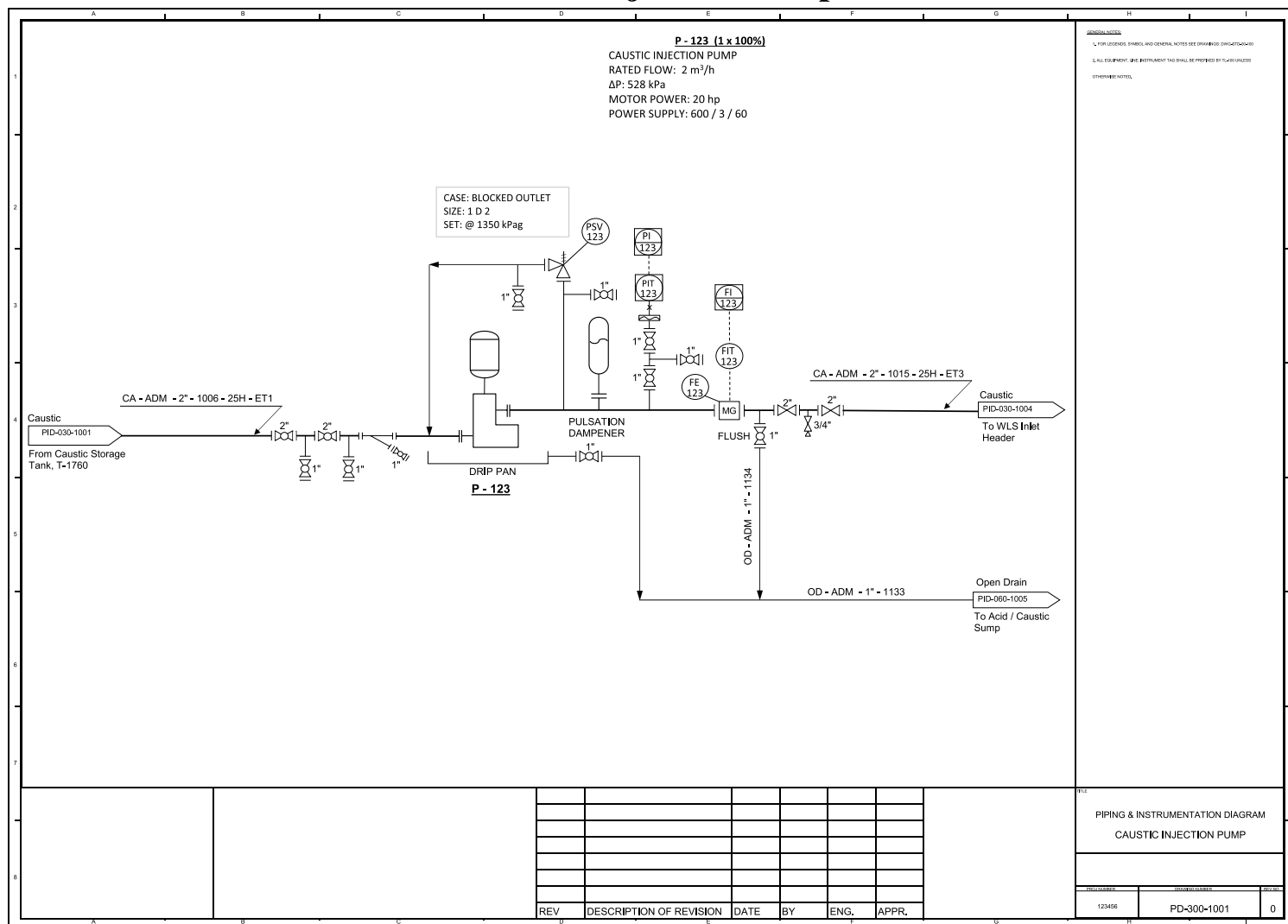
Plate Heat Exchanger



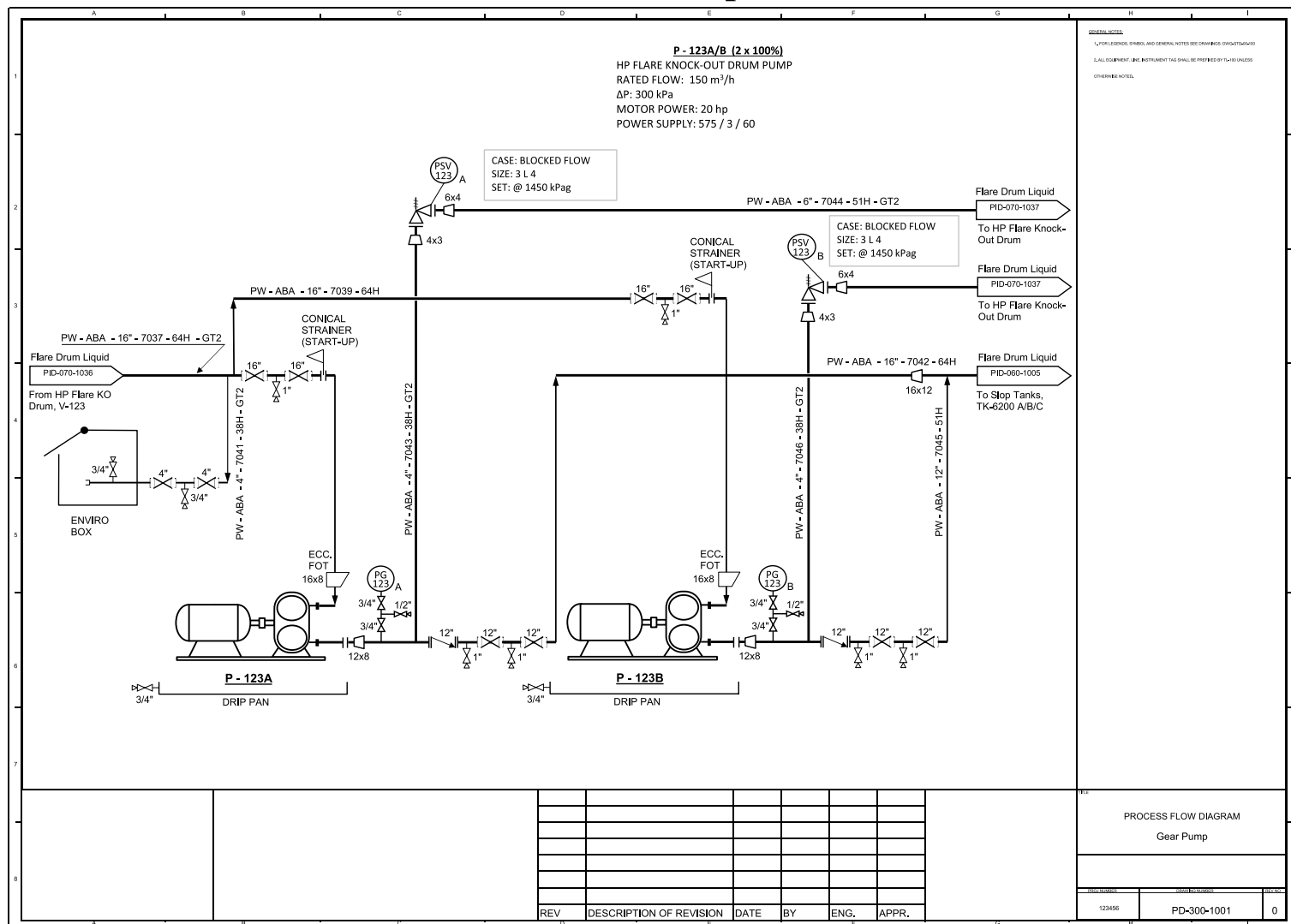




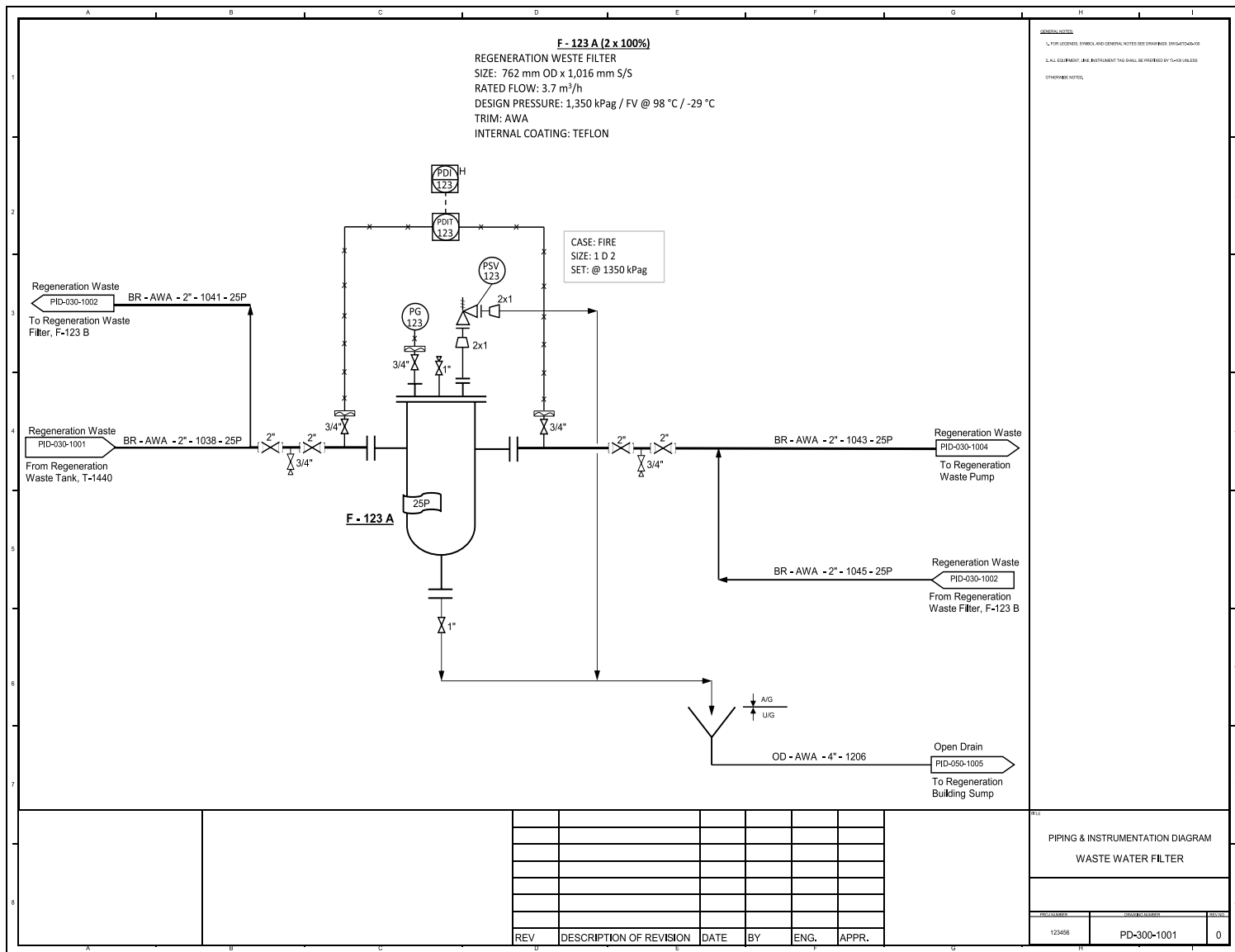




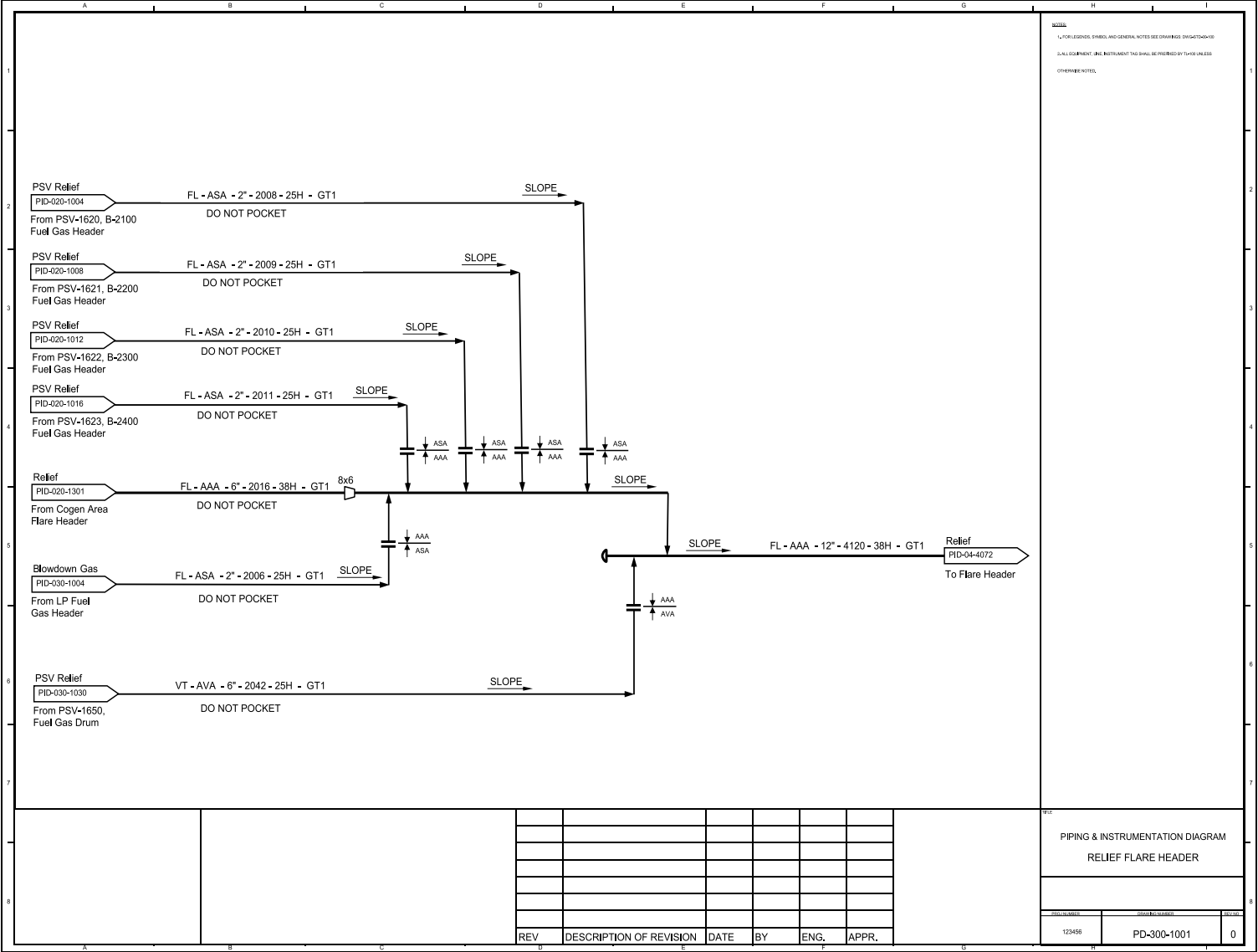
Gear Pump



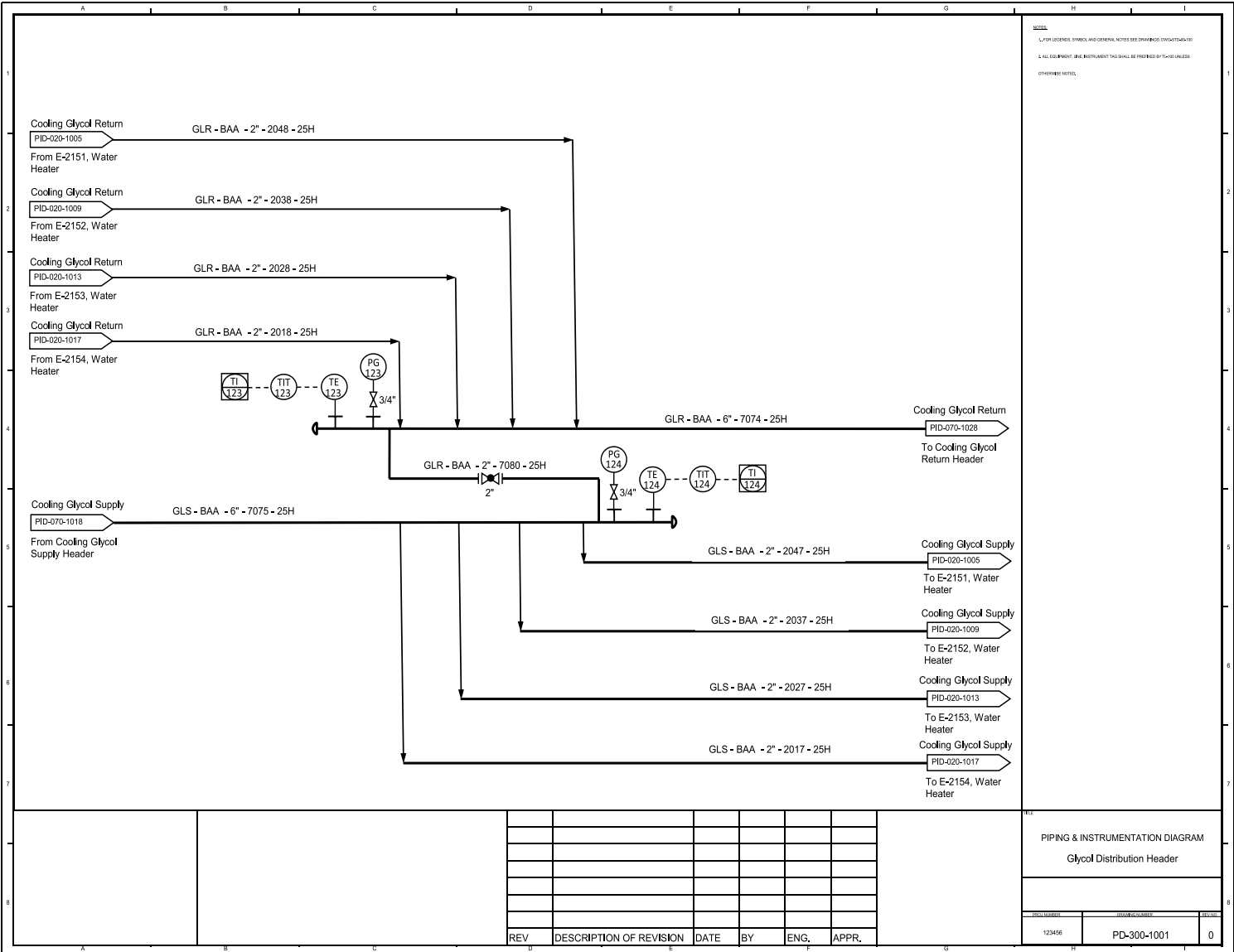
Waste Water Filter



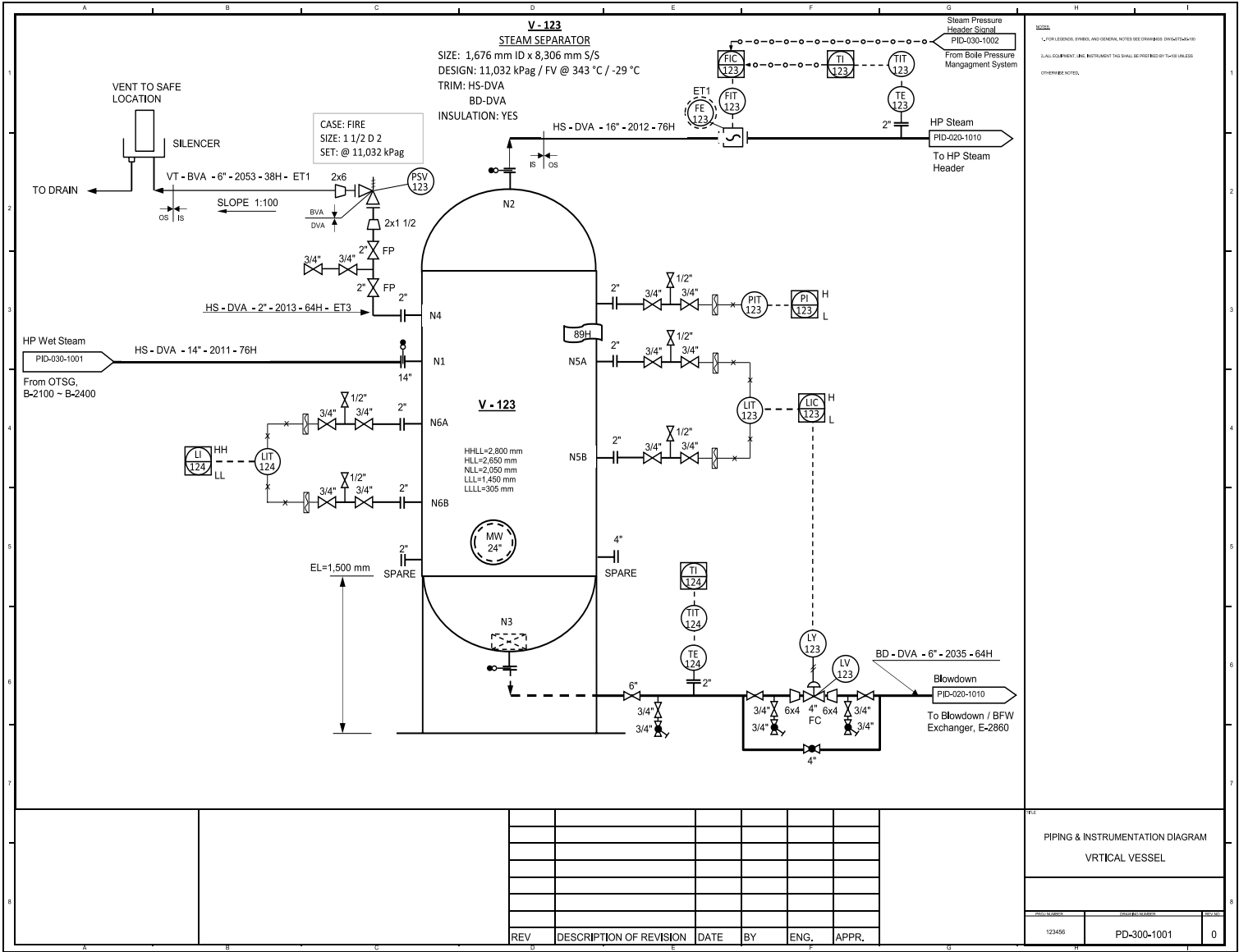
Relief Flare Header



Glyced Distribution Header



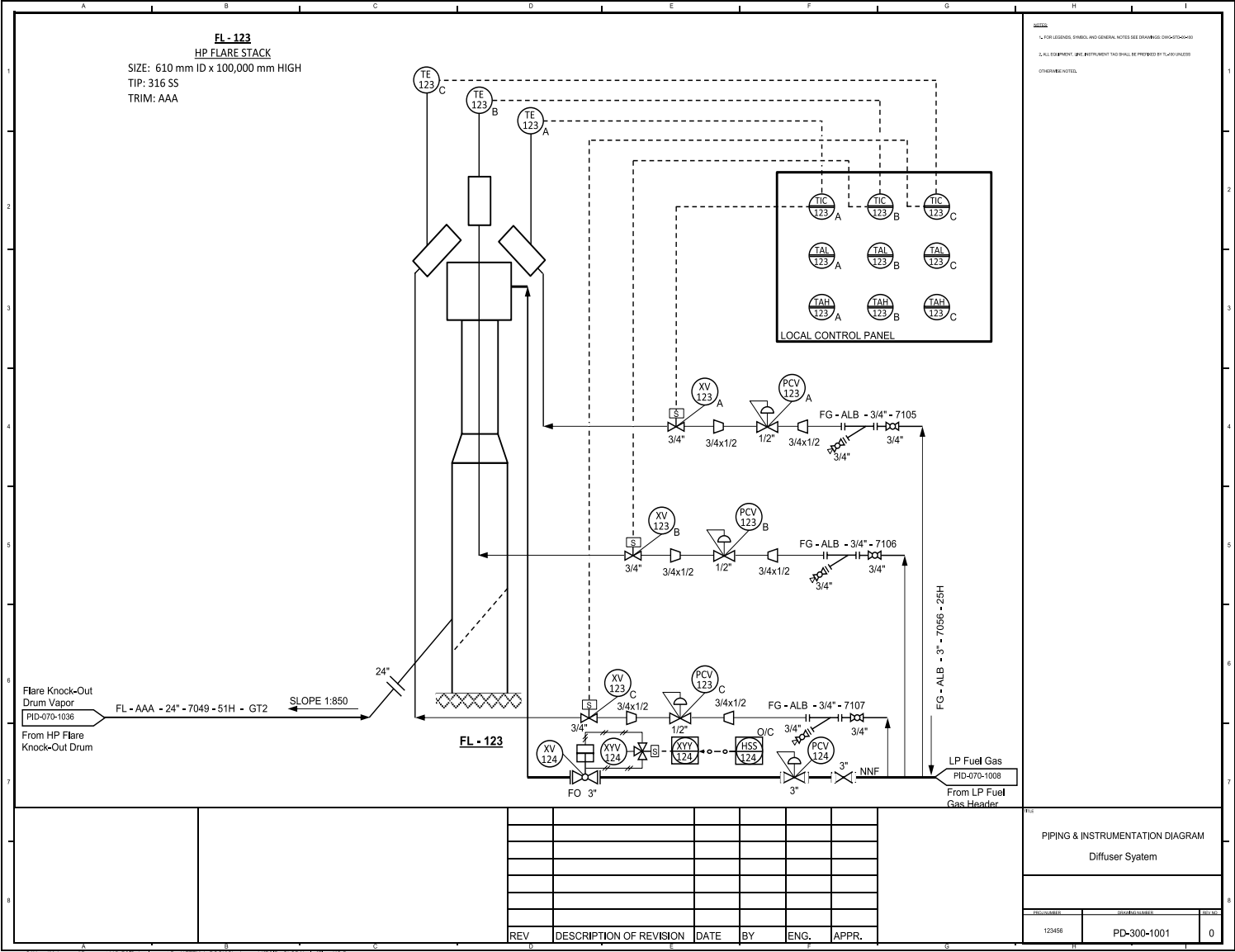
Vertical Vessel





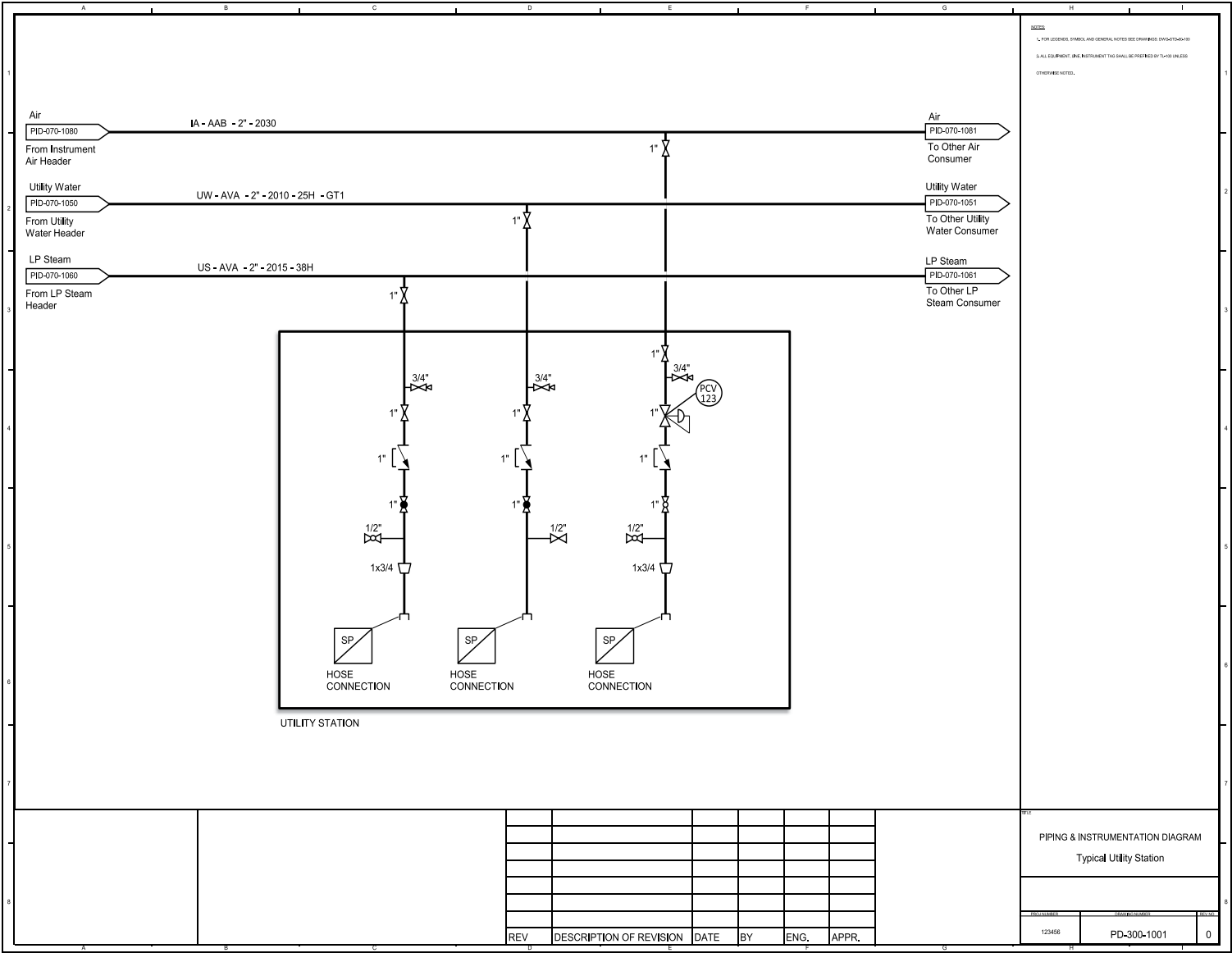


Diffuser System

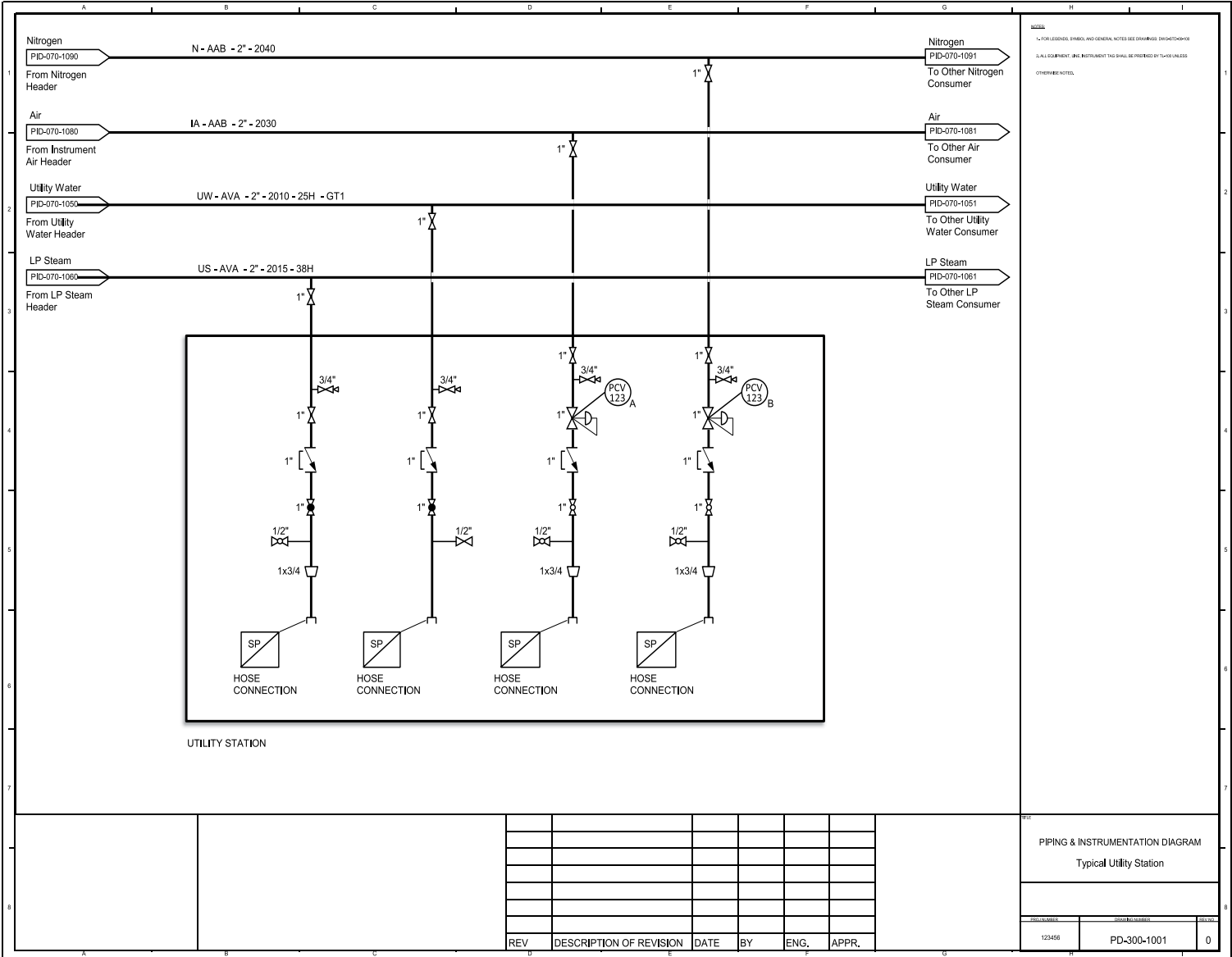




Typical Utility Station

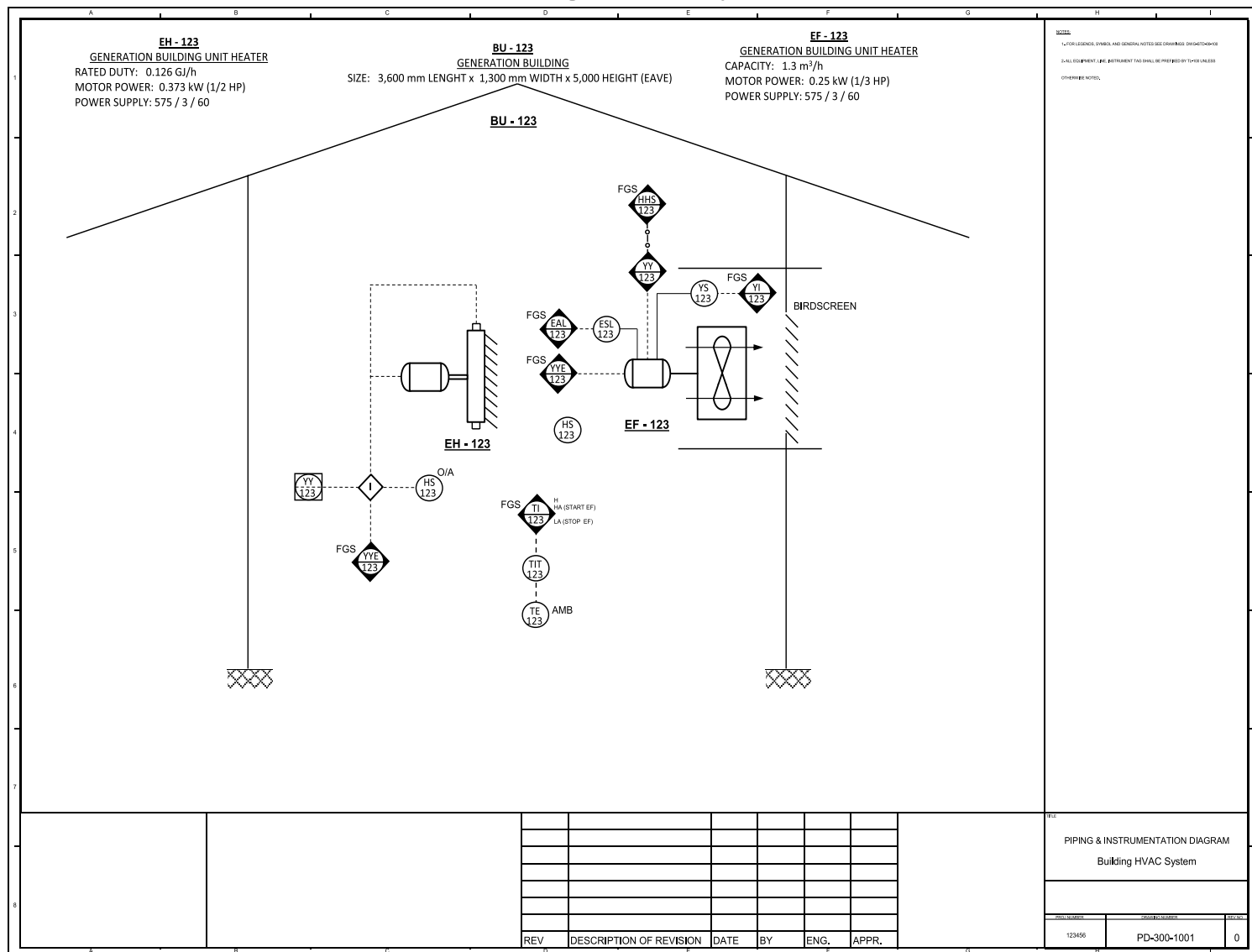


Typical Utility Station

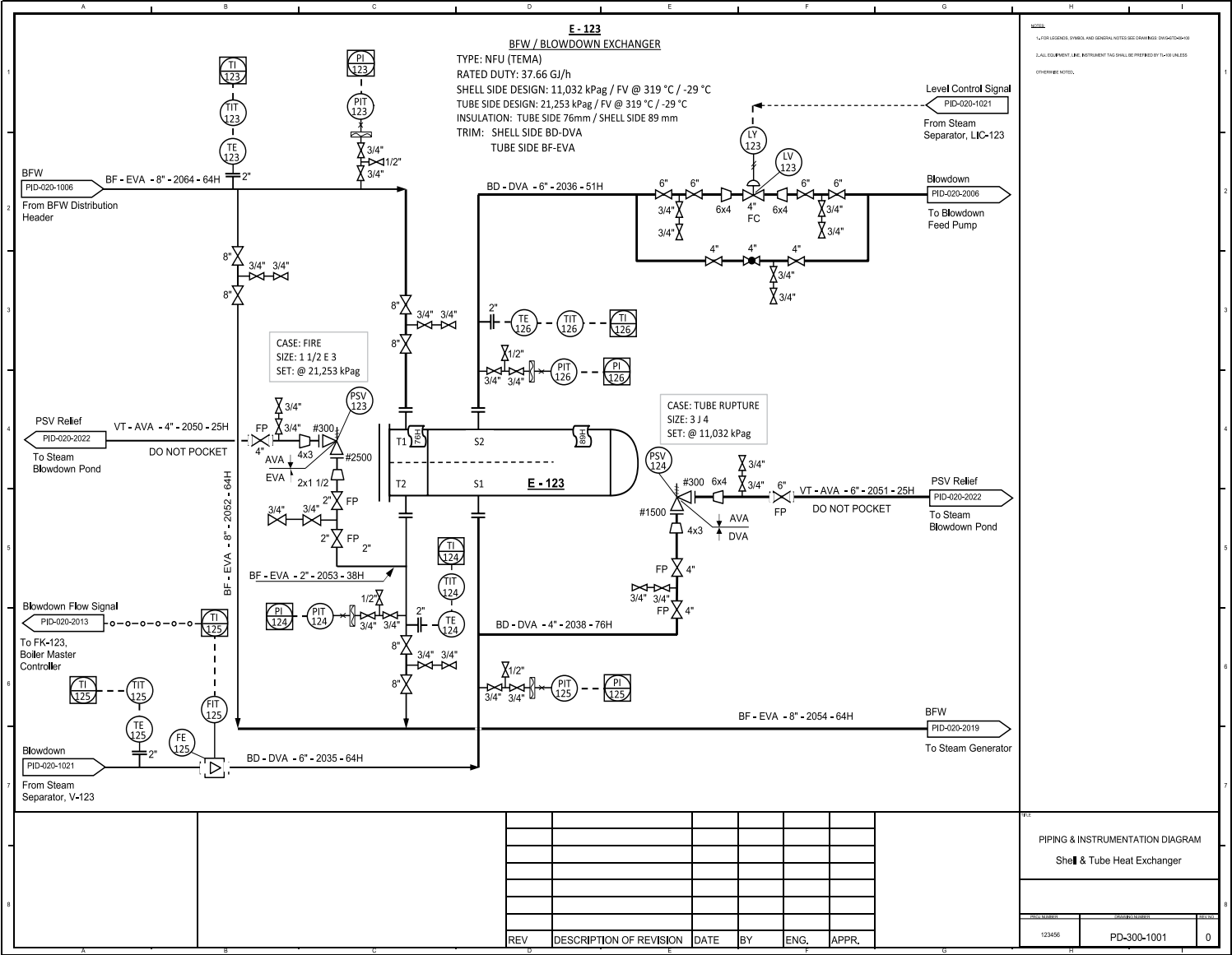


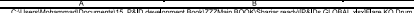
P and ID Development

Building HVAC System



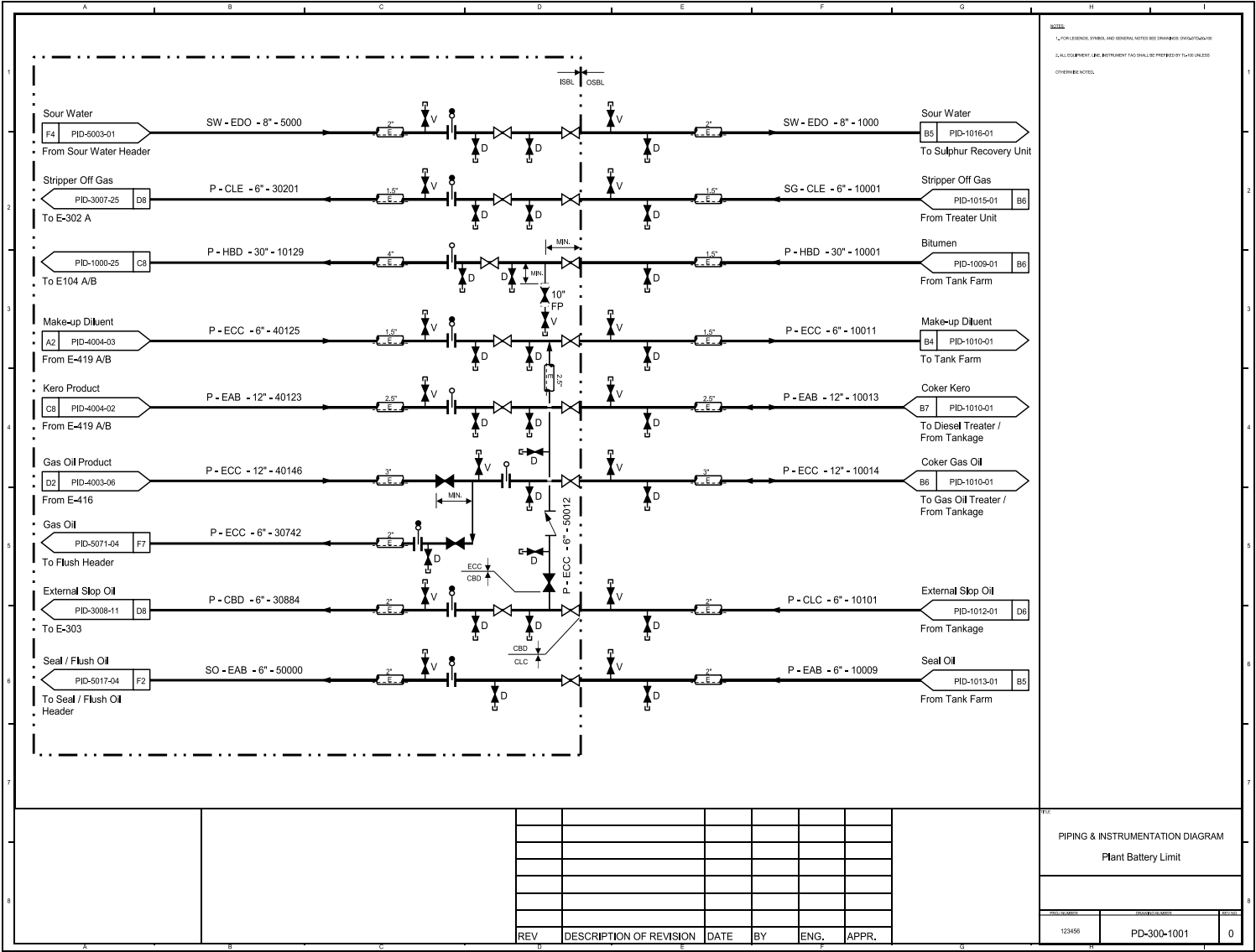
Shell & Tube Heat Exchanger



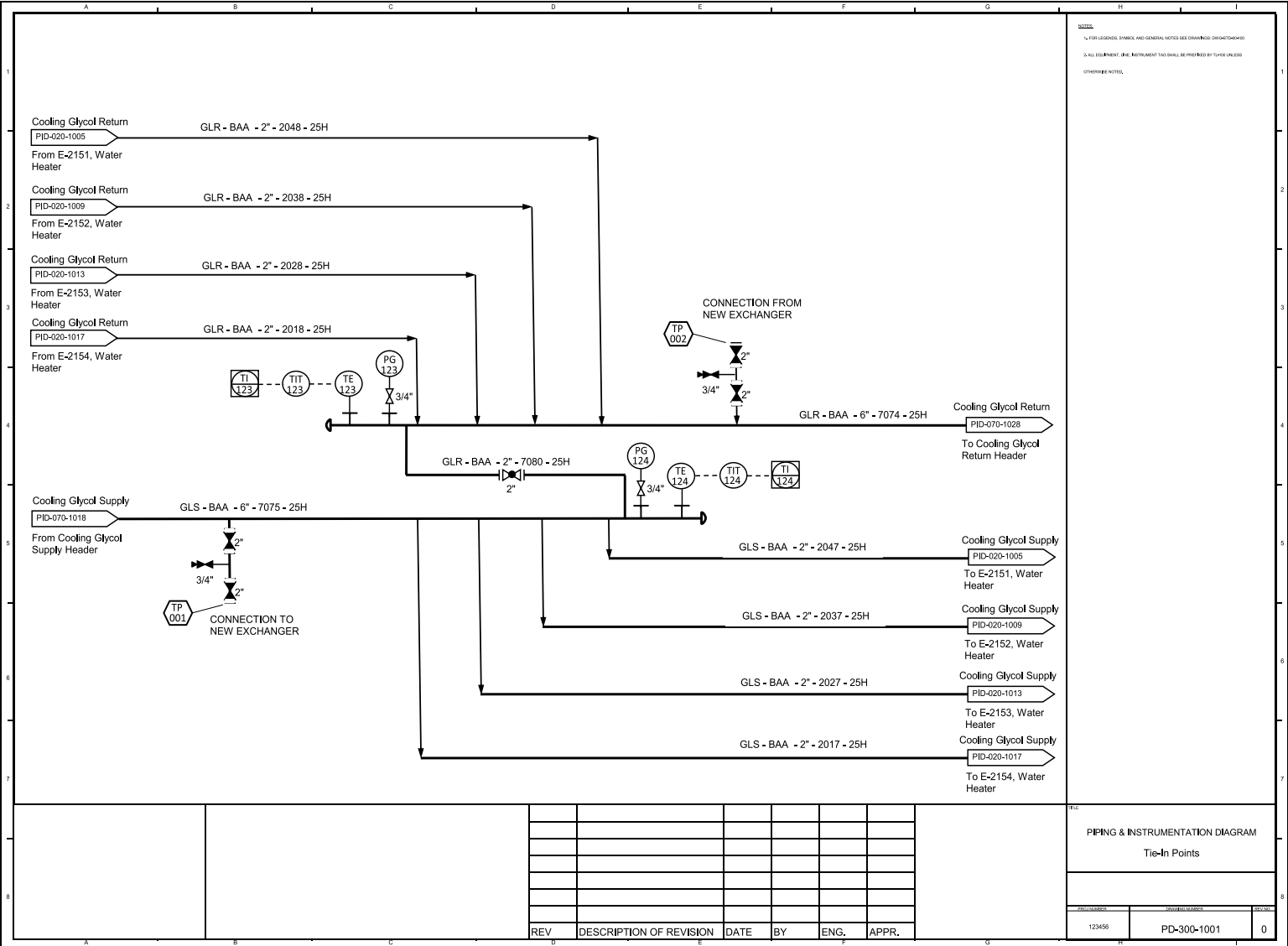




Plant Battery Limit

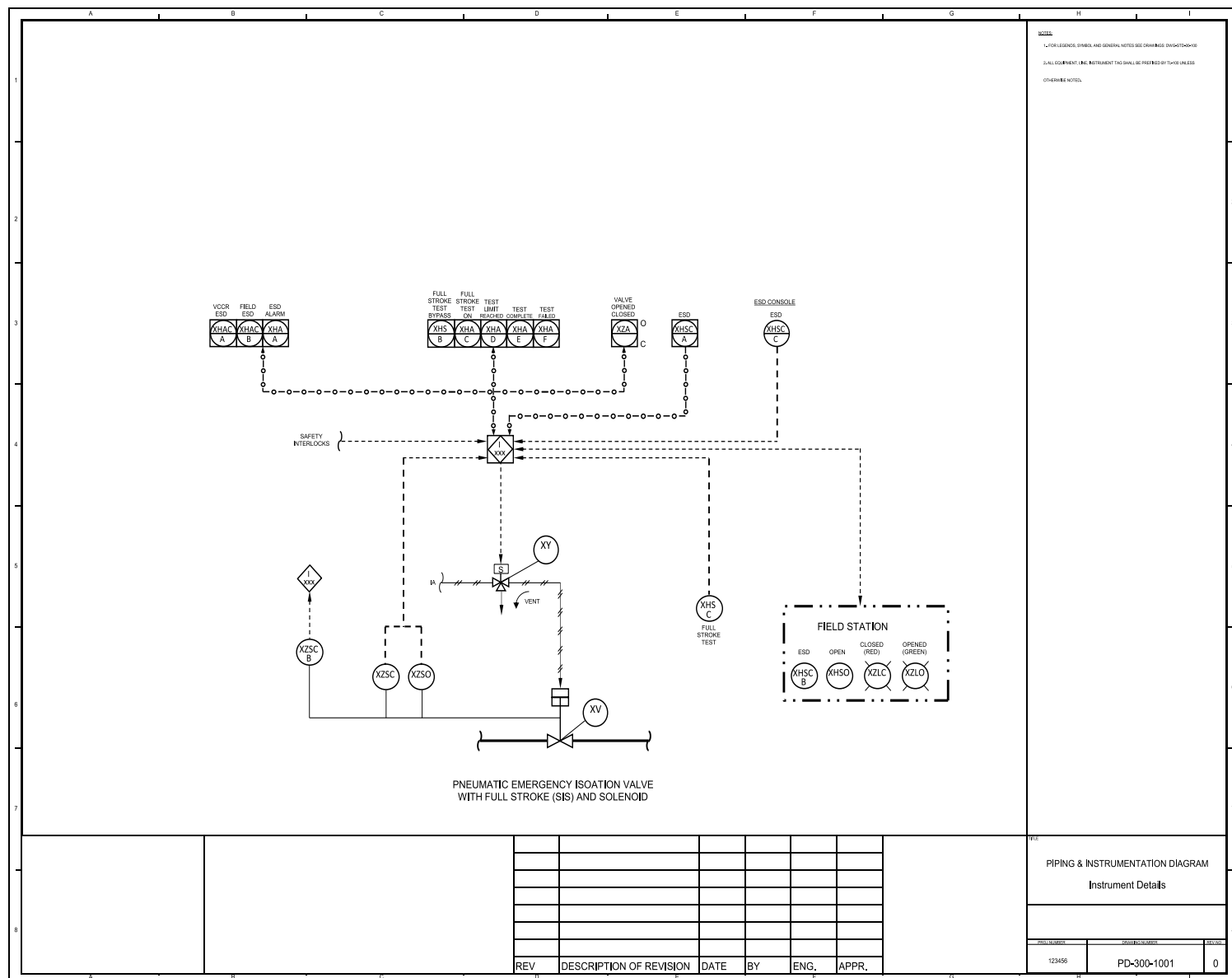


Tie - In Points



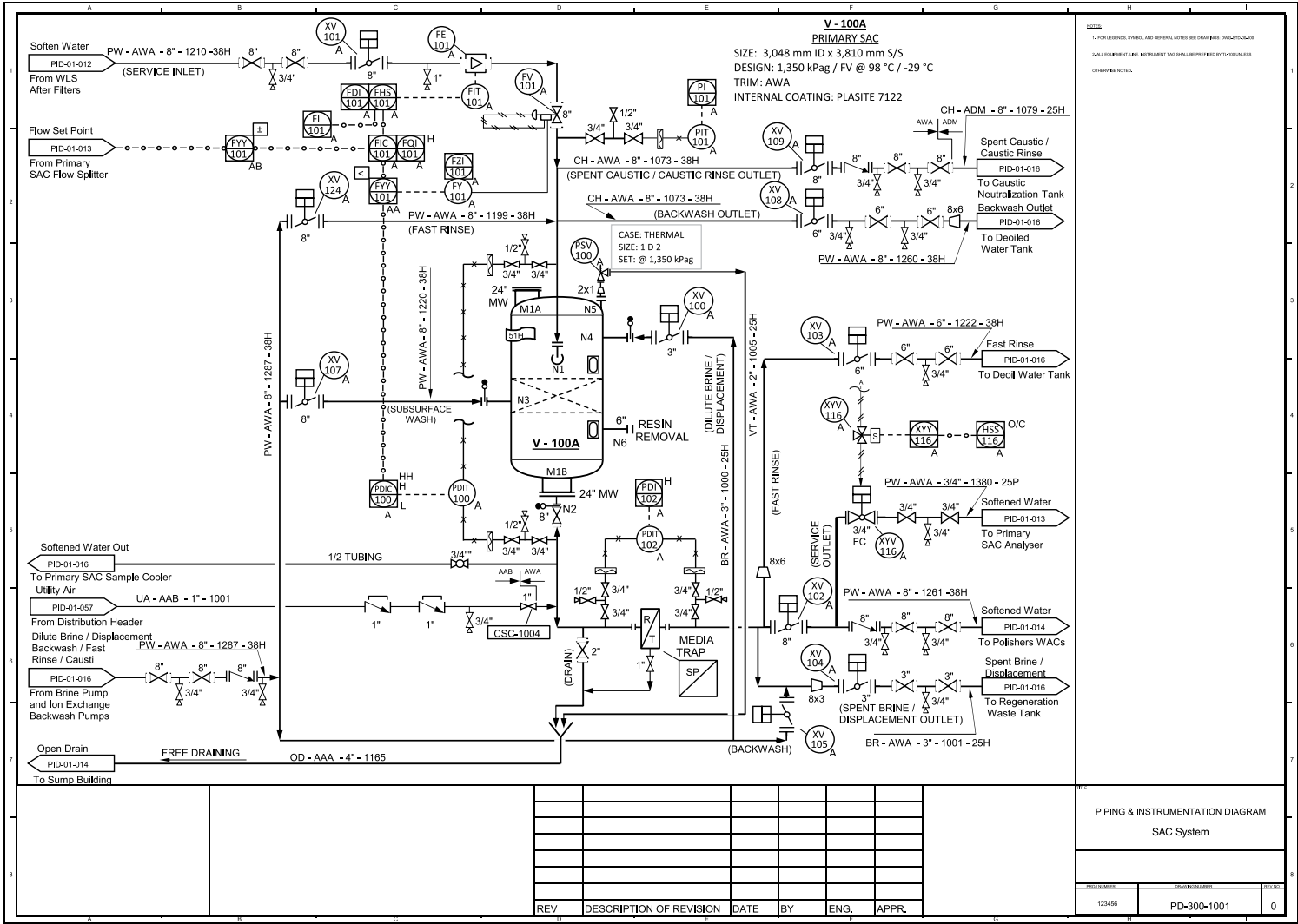


Instrument Details

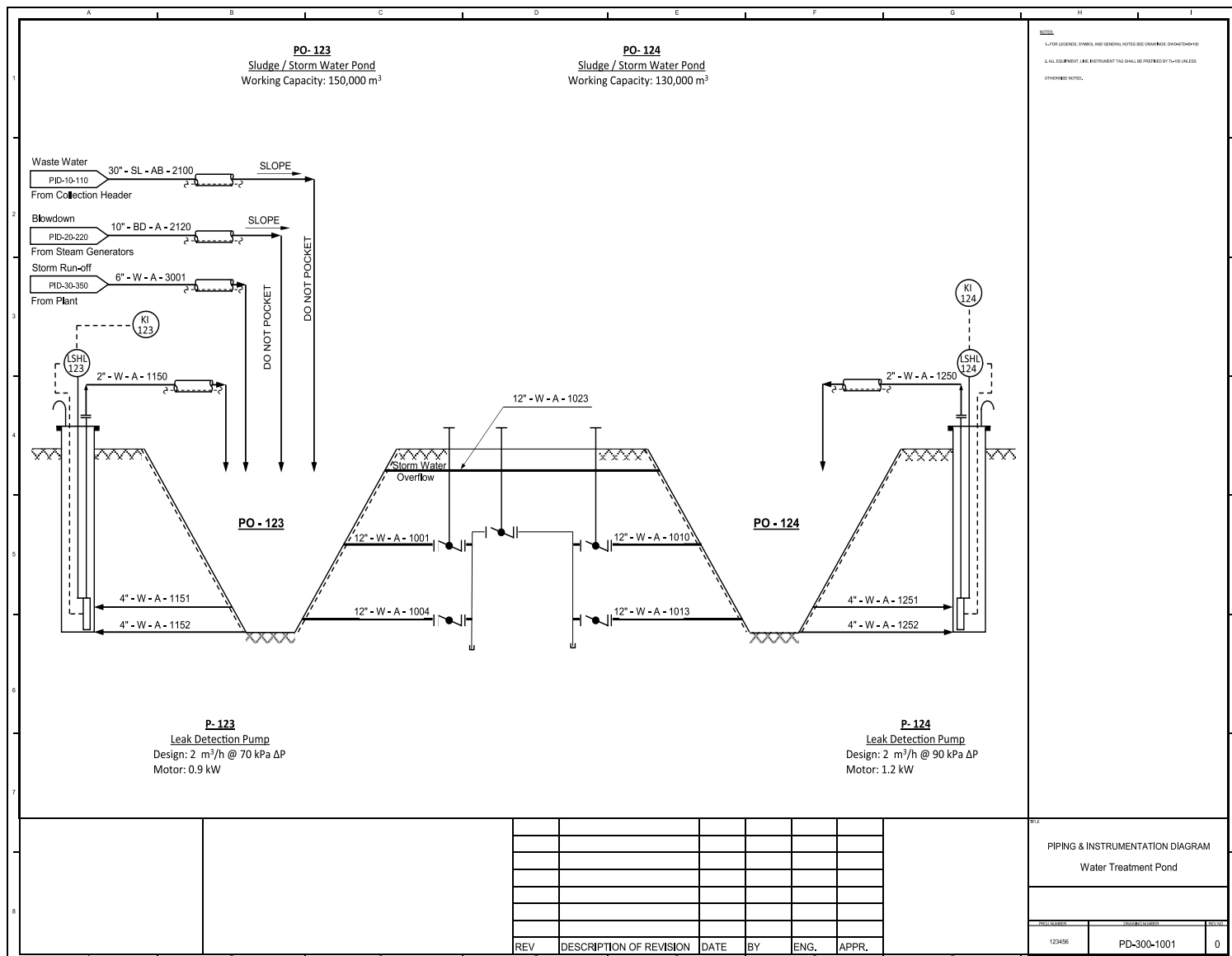




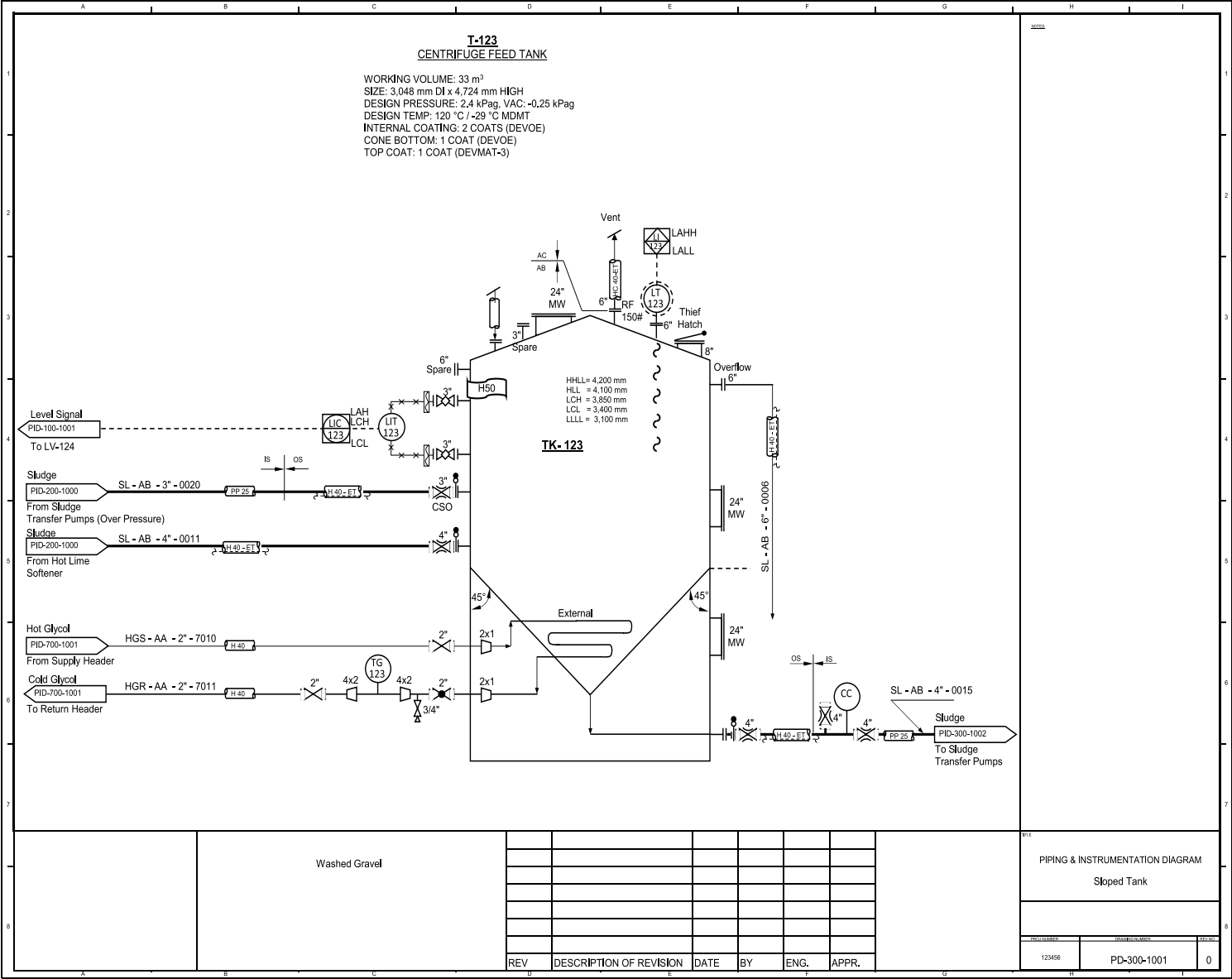
SAC System



Ponds



Sloped Tank







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