



VOLUME TWO

SURFACE PRODUCTION OPERATIONS



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Design of Gas-Handling Systems
and Facilities

3rd Edition

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PREFACE

As teachers of surface production facility design in petroleum engineering programs at Louisiana State University and the University of Houston, Ken and I realized that no single source could be used as a text in this field. We found ourselves reproducing pages from catalogs, reports, and projects we had done to provide our students with the basic information they needed to understand our lectures and carry out their assignments. More importantly, the material that did exist usually contained nomographs, charts, and rules of thumb that did not refer to the basic theories and underlying assumptions upon which they were based. Although **Volume 2** often builds upon information that was presented in *Surface Production Operations, Volume 1: Design of Oil-Handling Systems and Facilities*, it does present the basic concepts and techniques necessary to select, specify, and size gas-handling, conditioning, and processing equipment. After the initial release of the first two volumes, we realized that many topics needed to be covered in more depth. As a result, three additional volumes were added to the series, to be published at a later time. **Volume 3** covers the design of facility piping and pipeline systems, as well as relief and vent disposal systems; **Volume 4** covers rotating equipment, specifically pumps, compressors, and drivers; and **Volume 5** covers instrumentation, process control, safety systems, electrical installations in hazardous (classified) locations, and project and risk management.

Volume 2, which covers one semester's work or a two-week short course, focuses on the processes and equipment involved with gas-handling, conditioning, and processing facilities. The objective of gas-handling, conditioning, and processing facilities is to separate natural gas, condensate, or oil and water from a gas-producing well and condition these fluids for sale or disposal. Specific areas addressed by this volume include basic principles and fluid properties, process selection, heat transfer and heat exchanger equipment, hydrate prediction and prevention, condensate stabilization, compression, dehydration, acid gas treating, and gas processing. As was the case with Volume 1, this text covers topics that are common to both oil- and gas-handling production facilities, such as surface safety systems.

Throughout the text we have attempted to concentrate on what we perceive to be modern and common practices. We have personally been involved in the design and troubleshooting of such modern facilities

throughout the world, or we have people in our organizations who have done so. Undoubtedly, we are influenced by our own experience and prejudices. We apologize if we left something out or have expressed opinions about equipment types that differ from your own. We have learned much from our students' comments about such matters and would appreciate receiving your input for future revisions/editions.

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I would like to thank **Ken Arnold** for his insight, knowledge, and guidance in the preparation of *Surface Production Operations, Volume 1: Design of Oil-Handling Facilities* and *Volume 2: Design of Gas-Handling Facilities*. Without Ken's expert contributions, Volumes 1 and 2 would not have been possible. Ken and I are also indebted to the many companies and individuals who, over the past 30 years, have aided us by providing hours of argument and insight about the various topics covered in these volumes. Without their help, Volumes 1 and 2 would not have been possible. I offer a final thank you to Heri Wibowo for his drafting and coordinating efforts, as well as the abilities he showed in putting everything together for the third edition of Volume 2.



Overview of Gas-Handling, Conditioning, and Processing Facilities

Volume 1 of the *Surface Production Operations* series focuses on the processes and equipment involved with oil-handling facilities. **Volume 2** focuses on the processes and equipment involved with gas-handling, conditioning, and processing of natural gas for sale. **Volume 3** focuses on plant (facilities) piping and pipeline systems. It also covers relief, vent and flare systems. **Volume 4** focuses of rotating equipment, specifically pumps, compressors and drivers. **Volume 5** focuses on instrumentation, process control and safety systems. The objective of gas-handling, conditioning, and processing facilities is to separate natural gas, condensate, or oil and water from a gas-producing well and condition these fluids for sale or disposal.

Chapter 2 discusses basic principles and fluid properties. Heat transfer theory and heat exchanger equipment are discussed in **Chapters 3** and **4**. Hydrate prediction and prevention are discussed in **Chapter 5**. Condensate stabilization, the process of flashing the lighter hydrocarbons to gas and maximizing the recovery of natural gas liquids (NGLs), is the topic of **Chapter 6**. The subject of **Chapter 7** is methods of gas dehydration, and specific comments on maintenance, care, and troubleshooting of glycol dehydration systems are the topics of **Chapter 8**. Gas sweetening, the removal of acid gas compounds (primarily carbon dioxide and hydrogen sulfide) from natural gas, is discussed in **Chapter 10**. Gas processing to extract natural gas components, primarily ethane, propane, butane and heavier NGLs, from a natural gas stream is discussed in **Chapter 10**. Treating the condensate, or oil and water, after the initial separation from the natural gas is covered in Volume 1.

Figure 1.1 is a simple block diagram of a production facility that is primarily designed to handle gas wells. Each of the main blocks is described herein.

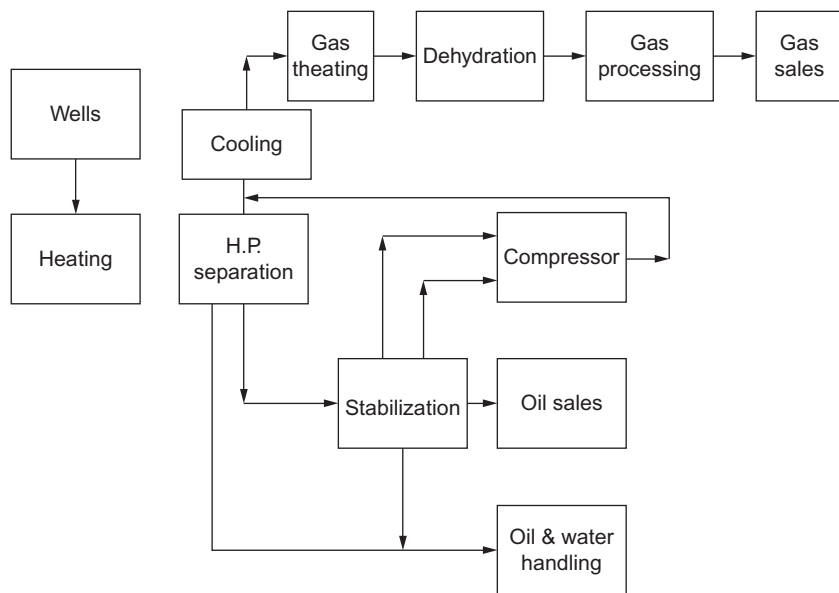


Figure 1.1 Gas field facility block diagram.



1.1 HEATING

The well flow stream may require heating prior to initial separation. Gas wells are often high pressure (HP) with shut-in tubing pressure of 5000–15,000 psig and flowing-tubing pressure (FTP) often exceeding 3000 psig. This pressure must be reduced to the appropriate operating pressure of the surface facilities and pipelines.

A choke is installed to control the flow. When the flow stream is choked, gas expands, temperature decreases, and liquids condense. If the temperature gets low enough, hydrates will form. Hydrates are crystalline solids made up of hydrocarbon and water molecules and form in the presence of hydrocarbon gas and liquid water at temperatures significantly higher than the freezing point of water. Hydrates can plug the flow-line, choke, and downstream equipment and reduce the capacity of vessels and pipelines. Therefore, flow from high-pressure gas wells may have to be heated to a temperature above the hydrate formation temperature before or immediately after the pressure drop from FTP to separator pressure, which occurs at the choke. Refer to [Chapter 5 on Hydrate Prediction and Prevention](#), of this volume, for a more complete discussion of hydrates.



1.2 SEPARATION

The HP separator provides a location for liquid to settle out from the gas. Normally, if the FTP is high enough, the separator pressure is set higher than the sales gas pipeline pressure so that the gas can go through the required cooling, treating, dehydration, and gas processing equipment, experience some pressure drop, and arrive at the required departing pipeline pressure.

Typically, in a gas facility, there is an initial separation at a HP, enabling reservoir energy to move the gas through the process to sales. It is very rare that the flowing-tubing pressure of a gas well, at least initially, is less than the gas sales pressure. With time, the flowing-tubing pressure may decline and compression may be needed prior to further handling of the gas. The initial separation is normally three-phase, as the separator size is dictated by gas capacity. That is, the separator will typically be large enough to provide sufficient liquid retention time for three-phase separation if it is to be large enough to provide sufficient gas capacity. Selection and sizing of separators are described in Volume 1.



1.3 COOLING

It is also possible that cooling may be necessary. If the flowing gas temperature is high, the temperature downstream of the choke may be high enough so that adding heat upstream of the HP separator is not required. If a substantial amount of gas and liquid is being produced, or if the flowing tubing temperature is very hot (e.g., from a deep high-pressure, high-temperature gas well), the gas temperature leaving the HP separator could be above the temperature that can easily be handled by the downstream processes. There is a maximum practical temperature for inlet streams to acid-gas treating and dehydration equipment. Hot gas leaving the HP separator carries more water vapor and increases the load on the downstream dehydration systems, which results in a much more expensive system than if the gas were cooled first. In addition, hot inlet gas could severely limit the ratio of discharge pressure to suction pressure that could be obtained from a downstream compressor. Thus, the gas may have to be cooled prior to compression, treating, and/or dehydration.

The cooler is typically an aerial cooler or a shell-and-tube exchanger that uses seawater or a contained cooling-water loop, which is cooled by

seawater or some other water source. Ideal gas temperatures are in the 80°–110 °F (27–43 °C) range.

In some fields, it may be necessary to provide heat during the early life of the wells when flowing-tubing pressures are high and there is a high-temperature drop across the choke. Later on, if the wells produce more liquid and the flowing-tubing pressure decreases, it may be necessary to cool the gas. Liquids retain the reservoir heat better and have less of a temperature drop associated with a given pressure drop than does gas.



1.4 STABILIZATION

Liquid from the initial separator is stabilized either by reducing the pressure and letting the lighter components flash (multistage flash separation) or by distillation, which is a combination of pressure reduction and heating (condensate stabilization). Most of the water will be removed during separation. The resulting stable condensate has a low vapor pressure so it can be stored in tanks for shipping at atmospheric pressure by truck, train, barge, or ship without excessive vapor venting. Often, there are vapor-pressure limitations that require liquid stabilization for pipeline shipments as well.

Stabilization of the hydrocarbon liquid maximizes the recovery of intermediate hydrocarbon components (C_3 to C_6) from the liquid stream. Multistage flash stabilization is discussed in Volume 1. Condensate stabilization is discussed in this volume in [Chapter 6](#).

The water and condensate removed in the separation/stabilization process must be separated, treated, and disposed of as described in Volume 1.



1.5 COMPRESSION

The lighter components removed in the stabilization process will be at a lower pressure than the HP gas stream. These components must be compressed to the HP separator pressure so that they can be recombined with the gas from the HP separator for further processing. Both reciprocating and centrifugal compressors are commonly used. In low-horsepower installations, especially for compressing gas from stock tanks (vapor recovery), screw or vane type compressors are common. Pumps, compressors, and drivers are covered in Volume 4.



1.6 GAS TREATING

Gas transmission companies require that impurities be removed from the gas they purchase. They recognize the need for removal of impurities for the efficient operation of their pipelines and for proper operation of their customers' gas-burning equipment. Consequently, contracts for the sale of gas to transmission companies always contain provisions regarding the quality of the gas that is delivered, and periodic tests are made to ensure those requirements are fulfilled by the seller.

Acid gases, usually hydrogen sulfide (H_2S) and carbon dioxide (CO_2), are impurities that are frequently found in natural gas and may have to be removed. Natural gas containing H_2S is called "sour gas." If the gas contains no H_2S , or if the H_2S has been removed, it is called "sweet gas." The process of removing the H_2S is referred to as "sweetening." The process of removing H_2S and CO_2 together is called "acid gas treatment," but sometimes the term "acid gas treatment" is used to denote removal of H_2S alone, CO_2 alone, or both.

Both H_2S and CO_2 can be very corrosive, with CO_2 forming carbonic acid, which is very corrosive in the presence of water. H_2S can potentially cause hydrogen embrittlement of steel. In addition, H_2S is extremely toxic at very low concentrations. When the gas is sold, the purchaser specifies the maximum allowable concentration of CO_2 and H_2S . A normal limit for CO_2 is between 2 and 4 volume percent, while H_2S is normally limited to **0.25 grain** per 100 standard cubic feet (scf) or 4 ppm by volume. Another common impurity of natural gas is nitrogen. Because nitrogen has essentially no calorific value, it lowers the heating value of gas. Gas purchasers may set a minimum limit of heating value (normally ~ 950 Btu/scf). In some cases, it may be necessary to remove the nitrogen to satisfy this requirement. This is done in very low temperature plants or with permeable membranes.

A common way to remove H_2S and CO_2 from natural gas is with an amine system, which uses a contact tower with trays or structured packing to pass the sour gas through the amine liquid, which absorbs the H_2S and some of the CO_2 . The amine is then regenerated in a stripping tower in which the H_2S and CO_2 are removed. There are several gas sweetening processes available to remove acid gases from natural gas. [Chapter 9](#) provides a further description of these processes.



1.7 GAS DEHYDRATION

Natural gas produced from a well is usually saturated with water vapor. Most gas treating processes also leave the gas saturated with water vapor. The water vapor itself is not objectionable, but the liquid or solid phase of water that may occur when the gas is compressed or cooled is troublesome. Liquid water accelerates corrosion of pipelines and other equipment; can allow the formation of hydrates that can plug valves, fittings, and sometimes the pipeline itself; and can accumulate in low points of pipeline, reducing the capacity of the lines. Removal of the water vapor by dehydration eliminates these possible difficulties and is normally required by gas sales agreements.

When gas is dehydrated its dew point (the temperature at which water will condense from the gas) is lowered. A typical dehydration specification in the U.S. Gulf Coast is 7 lb of water vapor per MMscf of gas (7 lb/MMscf). This gives a dew point of around 32 °F for 1000 psi gas. In the northern areas of the United States, Canada, Northern Europe, and Northern Asia the gas contracts require lower dew points or lower water vapor concentrations in the gas. Water vapor concentrations of 4 lb/MMscf are common. This gives a dew point of around 0 °F (−18 °C) for 1000 psi gas. If the gas is to be processed at very low temperatures, as in a cryogenic gas plant, water vapor removal down to 1 ppm may be required. Water is often removed from gas with a glycol dehydration system. Other methods include solid-bed desiccant adsorption, refrigeration, and membrane permeation. For a more detailed description of gas dehydration, refer to [Chapters 7 and 8](#).



1.8 GAS PROCESSING

Often the value received for gas depends on its heating value. However, if there is a market for ethane, propane, butane, and other gases, it may be economical to remove these components from the gas even though this will lower the heating value of the gas. In some cases, where the gas sales pipeline directly supplies a residential or commercial area with fuel, the sales contract may limit the heating value of the gas. The gas may then have to be processed to minimize its Btu content even if the extraction process is not economically justified.

Natural gas may be processed to recover liquid hydrocarbons in the form of NGLs, or liquefied petroleum gas (LPG). NGLs are hydrocarbon liquids, such as ethane, propane, butane, and natural gasoline, which can be separated from a natural gas stream after the heavier hydrocarbons components have been removed by separation at ambient temperatures. LPG is a mixture of hydrocarbons, principally butane and propane, which can be transported as a liquid under pressure, or at very low temperatures, and converted to gas on release of the pressure.

Chapter 10 discusses the most common processes used to separate NGL or LPG such as J-T valve, refrigeration, and cryogenic processes. The lean gas remaining can be used as fuel, reinjected into the reservoir, or put into a pipeline for sale.

A natural gas stream comprised of mostly methane can be cooled to very low temperatures and converted into a liquid called liquefied natural gas, LNG, to make it easy to transport by truck, rail, barge, or ship. Specialized processes and equipment are required for the process of liquefaction. The design and operation of LNG plants is beyond the scope of this book.

Throughout the processes in both oil and gas fields, care must be exercised to ensure that the equipment is capable of withstanding the maximum pressures to which it is subjected. Volume 3 discusses procedures for determining the line size and wall thickness of pipe, pressure ratings, and pressure (spec) breaks. In Volume 1, Chapter 6 Mechanical Design of Pressure Vessels discusses procedures for choosing the wall thickness of pressure vessels. The design pressure, or maximum allowable working pressure, of the piping system/process component is determined by a relief valve. Volume 3 includes chapters on pressure relief valves and relief, vent, and flare systems. Because safety considerations are so important in any facility design, Chapter 11 of this volume is devoted to safety analysis and safety system design.

Table 1.1 describes a gas field. The example problems that are worked in many of the sections of this text are for sizing the individual pieces of equipment needed for this field.

Table 1.1 Example Field

Q_g —Gas flow rate (total 10 wells)	100 MMscfd
SIBHP—Shut-in bottom-hole pressure	8000 psig
SITP—Shut-in tubing pressure	5000 psig
Initial FTP—Initial flowing-tubing pressure	4000 psig
Final FTP—Final flowing-tubing pressure	1000 psig
Initial FTT—Initial flowing-tubing temperature	120 °F
Final FTT—Final flowing-tubing temperature	175 °F
BHT—Bottom-hole temperature	224 °F

Separator Gas Composition (1000 psia)

Component	Mole %
CO ₂	4.03
N ₂	1.44
C ₁	85.55
C ₂	5.74
C ₃	1.79
<i>i</i> C ₄	0.41
<i>n</i> C ₄	0.41
<i>i</i> C ₅	0.20
<i>n</i> C ₅	0.13
C ₆	0.15
C ₇ ⁺	0.15
H ₂ S	19 ppm

For C₇⁺—mol. wt. = 147, P_c = 304 psia, T_c = 1112 °R.

Condensate—60 bbl/MMscf, 52.3 °API.

Initial free-water production—0 bbl/MMscf.

Final free-water production—15 bbl/MMscf (at surface conditions).

Gas sales requirements—1000 psi, 7 lb/MMscf, 1/4 grain H₂S, 2% CO₂.



Basic Principles



2.1 INTRODUCTION

Before discussing gas-handling, conditioning, and processing equipment and design techniques, it is necessary to review some basic principles and fluid properties. We will also discuss some of the common calculation procedures, conversions, and operations used to describe the fluids encountered in production operations.



2.2 FLUID ANALYSIS

An example fluid analysis of a typical gas well is shown in [Table 2.1](#). Note that only paraffin hydrocarbons are shown. This is not correct, even though only paraffin hydrocarbons may be the predominant series present. Also note that all molecules of heptane and larger ones are lumped together as heptanes plus fraction.

Sometimes operators request more complete analysis with many more components listed. Process engineers often like to input all of this data into their simulation models, but from a practical standpoint the exact composition of these higher-end components makes little difference in sizing equipment. In addition, fluid samples often are not totally representative of what is eventually produced through the facility as new wells are drilled and reservoir conditions change. Optimizing a design to a precise set of “given” fluid properties may result in a facility that is not flexible enough to handle the inevitable changes that occur in oil and gas field developments.



2.3 PHYSICAL PROPERTIES

An accurate estimate of physical properties is essential if one is to obtain reliable calculations. Physical and chemical properties depend on:

Table 2.1 Example Fluid Analysis of a Gas Well

Component	Mol%
Methane (C ₁)	35.78
Ethane (C ₂)	21.46
Propane (C ₃)	1.40
<i>i</i> -Butane (<i>i</i> -C ₄)	5.35
<i>n</i> -Butane (<i>n</i> -C ₄)	10.71
<i>i</i> -Pentane (<i>i</i> -C ₅)	3.81
<i>n</i> -Pentane (<i>n</i> -C ₅)	3.07
Hexanes (C ₆)	3.32
Heptanes plus (C ₇₊)	3.24
Nitrogen	0.20
Carbon dioxide	1.66
Total	100.00

- Pressure
- Temperature
- Composition

Most hydrocarbon streams are mixtures of hydrocarbons that may contain varying quantities of contaminants such as:

- Hydrogen sulfide
- Carbon dioxide
- Water

The more similar the character of the mixture molecules, the more orderly is their behavior. A single-component system composed entirely of a simple molecule, such as methane, behaves in a very predictable, correctable manner.

The accuracy of calculations decrease in the following order:

- Single-component system
- Mixture of molecules from the same homologous series
- Mixture of molecules from different homologous series
- Hydrocarbon mixtures containing sulfur compounds and/or carbon dioxide

Performance data for a single-component system can be accurately correlated in graphic or tabular form. For all others, one must use either pressure/volume/temperature (*PVT*) equations of state or the weighted average. The weighted average assumes that the contribution of an individual molecule is in proportion to its relative quantity in the mixture. The more dissimilar the molecules, the less accurate the prediction becomes. [Table 2.2](#) lists some of the physical properties of some of the paraffin hydrocarbon series.

Water in liquid or vapor form is present to some degree in all systems. Liquid water is essentially immiscible in hydrocarbon liquids. However,

Table 2.2 Physical Properties of Paraffin Hydrocarbons

Component	Methane	Ethane	Propane	iso-Butane	n-Butane	iso-Pentane	n-Pentane	n-Hexane	n-Heptane	n-Octane	n-Nonane	n-Decane
Molecular weight	16.043	30.070	44.097	58.124	58.124	72.151	72.151	86.178	100.205	114.232	128.259	142.286
Boiling point @ 14.696 psia, °F	-258.73	-127.49	-43.75	10.78	31.08	82.12	96.92	155.72	209.16	258.21	303.47	345.48
Freezing point @ 14.696 psia, °F	-296.44	-297.49	-305.73	-255.28	-217.05	-255.82	-201.51	-139.58	-131.05	-70.18	-64.28	-21.36
Vapor pressure @ 100 °F, psia	(5000.0)	(800.0)	188.4	72.58	51.71	20.445	15.574	4.960	1.620	0.5369	0.1795	0.0609
Density of liquid @ 60°F and 14.696 psia												
Relative density @ 60 °F/60 °F	(0.3)	0.3562	0.5070	0.5629	0.5840	0.6247	0.6311	0.6638	0.6882	0.7070	0.7219	0.7342
°API	(340.0)	265.6	147.3	119.8	110.7	95.1	92.7	81.60	74.08	68.64	64.51	61.23
Absolute density, lbm/gal (in vacuum)	(2.5)	2.970	4.227	4.693	4.870	5.208	5.262	5.534	5.738	5.894	6.018	6.121
Apparent density, lbm/gal (in air)	(2.5)	2.960	4.217	4.683	4.861	5.198	5.252	5.524	5.729	5.885	6.008	6.112

Continued

Table 2.2 Physical Properties of Paraffin Hydrocarbons—cont'd

Component	Methane	Ethane	Propane	iso- Butane	n- Butane	iso- Pentane	n- Pentane	n- Hexane	n- Heptane	n- Octane	n- Nonane	n- Decane
Density of gas @ 60 °F and 14.696 psia												
Relative density (air = 1), ideal gas	0.5539	1.0382	1.5225	2.0068	2.0068	2.4911	2.4911	2.9755	3.4598	3.9441	4.4284	4.9127
lb/M ft ³ , ideal gas	42.28	79.24	116.20	153.16	153.16	190.13	190.13	227.09	264.06	301.02	337.98	374.95
Volume @ 60 °F and 14.696 psia												
Liquid, gal/lb-mol	(6.4)	10.13	10.43	12.39	11.94	13.85	13.72	15.57	17.46	19.38	21.31	23.45
ft ³ has/gal	(59.1)	37.48	36.375	30.64	31.79	27.39	27.67	24.37	21.73	19.58	17.81	16.33
liquid, ideal gas												
Ratio, gas/liquid, in vacuum	(442.0)	280.4	272.1	229.2	237.8	204.9	207.0	182.3	162.6	146.5	133.2	122.2
Critical conditions												
Temperature, °F	−116.67	89.92	206.06	274.46	305.62	369.10	385.8	453.6	512.7	564.22	610.68	652.0
Pressure, psia	666.4	706.5	616.0	527.9	550.6	490.4	488.6	436.9	396.8	360.7	331.8	305.2
Gross calorific value, combustion @ 60 °F												
Btu/lb, liquid	−	22,181	21,489	21,079	21,136	20,891	20,923	20,783	20,679	20,607	20,543	20,494
Btu/lb, gas	23,891	22,332	21,653	21,231	21,299	21,043	21,085	20,942	20,838	20,759	20,700	20,651

Btu/ft ³ , ideal gas	1016.0	1769.6	2516.1	3251.9	3262.3	4000.9	4008.9	4755.9	5502.5	6248.9	6996.5	7742.9
Btu/gal, liquid	-	65,869	90,830	98,917	102,911	108,805	110,091	115,021	118,648	121,422	123,634	125,448
Volume air to burn one volume, ideal gas	9.54	16.71	23.87	31.03	31.03	38.19	38.19	45.35	52.52	59.68	66.84	74.00
Flammability limits @ 100 °F and 14.696 psia												
Lower, volume % in air	5.0	2.9	2.0	1.8	1.5	1.3	1.4	1.1	1.0	0.8	0.7	0.7
Upper, volume % in air	15.0	13.0	9.5	8.5	9.0	8.0	8.3	1.7	7.0	6.5	5.6	5.4
Heat of vaporation @ 14.696 psia												
Btu/lb @ boiling point	219.45	211.14	183.01	157.23	165.93	147.12	153.57	143.94	163.00	129.52	124.36	119.65
Specific heat @ 60 °F and 14.696 psia												
C _p gas, Btu/ (lb-°F), ideal gas	0.5267	0.4078	0.3885	0.3867	0.3950	0.3844	0.3882	0.3863	0.3845	0.3833	0.3825	0.3818
C _v gas, Btu/ (lb-°F), ideal gas	0.4029	0.3418	0.3435	0.3525	0.3608	0.3869	0.3607	0.3633	0.3647	0.3659	0.3670	0.3678
K = C _p /C _v , ideal gas	1.307	1.193	1.131	1.097	1.095	1.077	1.076	1.064	1.054	1.048	1.042	1.038
C _p liquid, Btu/(lb-°F)	-	0.9723	0.6200	0.5707	0.5727	0.5333	0.5436	0.5333	0.5280	0.5241	0.5224	0.5210

depending on pressure and temperature, some water vapor will be present in the hydrocarbon gas phase. Because normal phase behavior calculations do not apply for water, special procedures must be used.

Equations of state use the values of P , V , and T at the critical point. Each molecular species has a unique critical point.

For each of the pure components shown in the tables, the critical values represent the maximum pressure and temperature at which a two-phase, vapor-liquid system can exist. Above P_C and T_C , only a single phase is possible. For mixtures, pseudocritical values are calculated that are correlation constants only and are not a point on the phase diagram.

2.3.1 Equations of State

The correlations that follow are commonly used for hydrocarbon systems and are suitable for most calculations. Any equation correlating P , V , and T is called an equation of state. The ideal equation of state is sometimes called ideal gas law, perfect gas law, or general gas law and is expressed by Equation (2.1).

$$PV = nRT \tag{2.1}$$

Where P = absolute pressure; V = volume; n = number of moles of gas of volume V at P and T ; R =universal gas constant (refer to Table 2.3); T =absolute temperature.

Equation (2.1) is valid up to pressures of about 60 psia (500 kPa, 4 bara). As pressure increases above this level, its accuracy decreases and the system should be considered a nonideal gas. Table 2.3 lists the values of the universal gas constant for different unit systems.

2.3.2 Molecular Weight and Apparent Molecular Weight

The number of moles is defined as follows:

$$\text{Mole} = \frac{\text{Mass}}{\text{Molecular weight}} \tag{2.2}$$

Table 2.3 Universal Gas Constant

P	V	T	R
kPa	m ³	K	8.314 (kPa)(m ³)/(kmol)(K)
MPa	m ³	K	0.00831 (MPa)(m ³)/(kmol)(K)
bar	m ³	K	0.08314 (bar)(m ³)/(kmol)(K)
psi	ft ³	°R	10.73 (psia)(ft ³)/(lb-mol)(°R)
lb/ft ²	ft ³	°R	1545 (psia)(ft ³)/(lb-mol)(°R)

expressed as

$$n = \frac{m}{M} \tag{2.3}$$

or in units as

$$\text{lb} - \text{mole}(\text{g} - \text{mole}) = \frac{\text{Mass in lb(g)}}{\text{Molecular Weight} \frac{\text{lb}}{\text{mole}} \left(\frac{\text{g}}{\text{mole}} \right)} \tag{2.4}$$

Molecular weight is defined as the sum of the atomic weights of the various elements present.

Example 2.1: Molecular Weight Calculation

Given

Determine the molecular weight of ethane, C_2H_6

Solution

Element	No. of Atoms		Atomic Weight		Product
C	2	×	12	=	24
H	6	×	1	=	6
Molecular weight					30 lb/mol

Up to now, we have addressed only pure substances. We now have to consider hydrocarbon mixtures. However, we must first discuss apparent molecular weight and specific gravity. It is not correct to say that a hydrocarbon mixture has a molecular weight; rather, it is an apparent molecular weight. Apparent molecular weight is defined as the sum of the products of the mole fractions of each component times the molecular weight of that component. This is shown in Equation (2.5):

$$\text{MW} = \sum \gamma_i (\text{MW})_i \tag{2.5}$$

where γ_i = molecular fraction of i th component; MW_i = molecular weight of i th component; $\sum \gamma_i = 1$.

Now, let us look at an example of the application of apparent molecular weight that will also result with a number that we will often use throughout this book.

Example 2.2: Determine the Apparent Molecular Weight of Dry Air, Which Is a Gas Mixture Consisting of Nitrogen, Oxygen, and Small Amounts of Argon

Given

Determine the apparent molecular weight of air given its approximate composition.

Gas Composition

Component	Mole Fraction
Nitrogen	0.78
Oxygen	0.21
Argon	0.01
Total	1.00

Solution

1. Look up the molecular weight of each component from the physical constant table.

$$(MW)_N = 28; (MW)_O = 32; (MW)_A = 40$$

2. Multiply the mole fraction of each component by its molecular weight.

$$\begin{aligned} (MW)_{AIR} &= \sum y_i (MW)_i = y_N (MW)_N + y_O (MW)_O + y_A (MW)_A \\ &= (0.78 \times 28) + (0.21 \times 32) + (0.01 \times 40) = \frac{29 \text{ lb}}{\text{lb/mole}} \end{aligned}$$

We will now define the specific gravity of a gas.

2.3.3 Gas Specific Gravity

The specific gravity of a gas is the ratio of the density of the gas to the density of air standard conditions of temperature and pressure.

Where

$$S = \frac{\rho_g}{\rho_{air}} \quad (2.6)$$

where ρ_g = density of gas; ρ_{air} = density of air.

Both densities must be computed at the same pressure and temperature, usually at standard conditions.

It may be related to the molecular weight by Equation (2.7).

$$S = \frac{(\text{MW})_g}{29} \quad (2.7)$$

Example 2.3: Calculate the Specific Gravity of a Natural Gas with the following composition
Given

Component	Mole Fraction (y_i)
Methane (C_1)	0.85
Ethane (C_2)	0.09
Propane (C_3)	0.04
<i>n</i> -Butane ($n-C_4$)	0.02
	1.00

2.3.4 Nonideal Gas Equations of State

The ideal gas equations of state describe most real gases at low pressure but do not yield reasonable results at higher pressures. Many PVT equations have been developed to describe nonideal, real gas behavior. Each is empirical in that it correlates a specific set of data using one or more empirical constants. Unfortunately, there is no correlation that is equally good for all gas mixtures. There can be as many such equations as there are individuals who correlate data. In some instances, the equations have been extrapolated beyond the compositions on which they were determined. This results in an inherent loss of accuracy.

The ideal equations of state can be approximated to the compressibility equation of state by multiplying the “ nRT ” part of the equation by Z :

$$PV = ZnRT \quad (2.8)$$

Where

$$Z = \frac{\text{Actual gas volume}}{\text{Ideal gas volume}} \quad (2.9)$$

If the gas acted as if it were an ideal gas, then the “ Z ” factor would be 1.0. The typical range of Z for most natural gasses is 0.8-1.2

The compressibility factor for a natural gas can be approximated from [Figures 2.1–2.6](#), which are from the **Engineering Data Book** of the Gas Processor Suppliers Association.

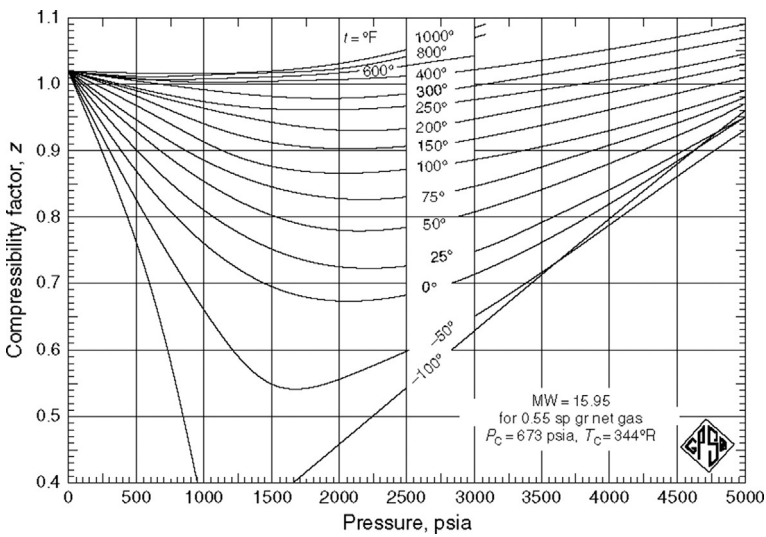


Figure 2.1 Compressibility factor for specific gravity. 0.55 gases. *Courtesy of GPSA Engineering Data Book.*

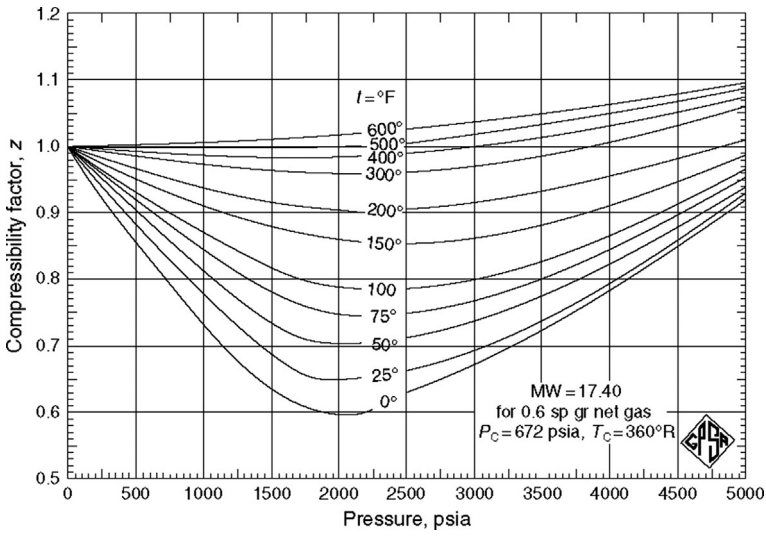


Figure 2.2 Compressibility factor for specific gravity. 0.6 gases. *Courtesy of GPSA Engineering Data Book.*

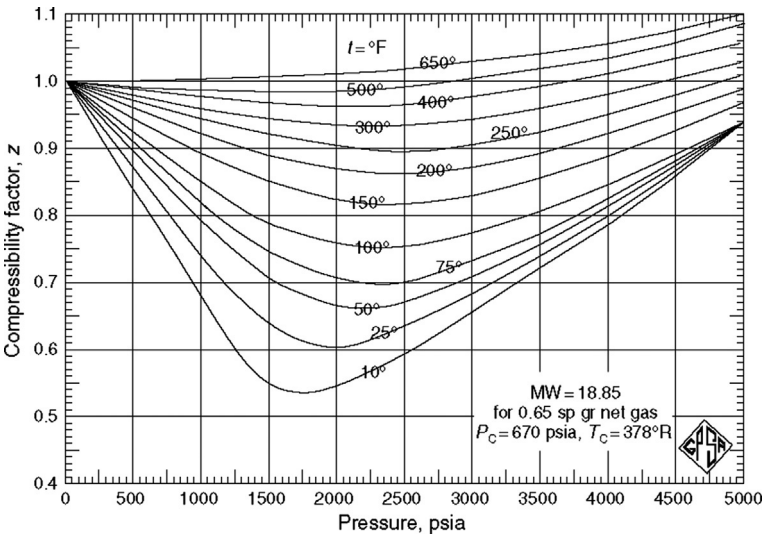


Figure 2.3 Compressibility factor for specific gravity. 0.65 gases. Courtesy of GPA Engineering Data Book.

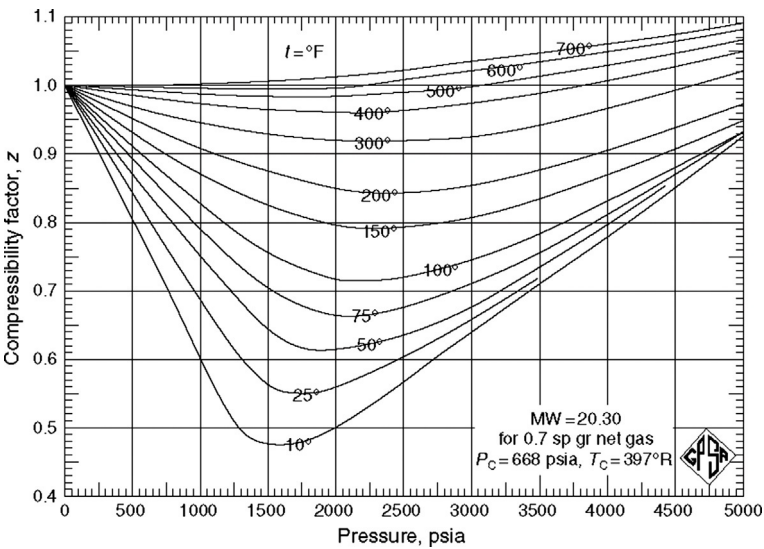


Figure 2.4 Compressibility factor for specific gravity. 0.7 gases. Courtesy of GPA Engineering Data Book.

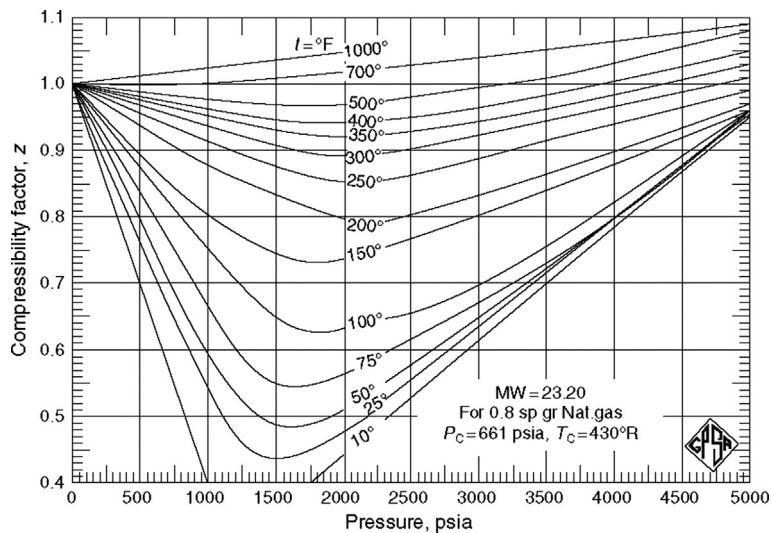


Figure 2.5 Compressibility factor for specific gravity. 0.8 gases. *Courtesy of GPSA Engineering Data Book.*

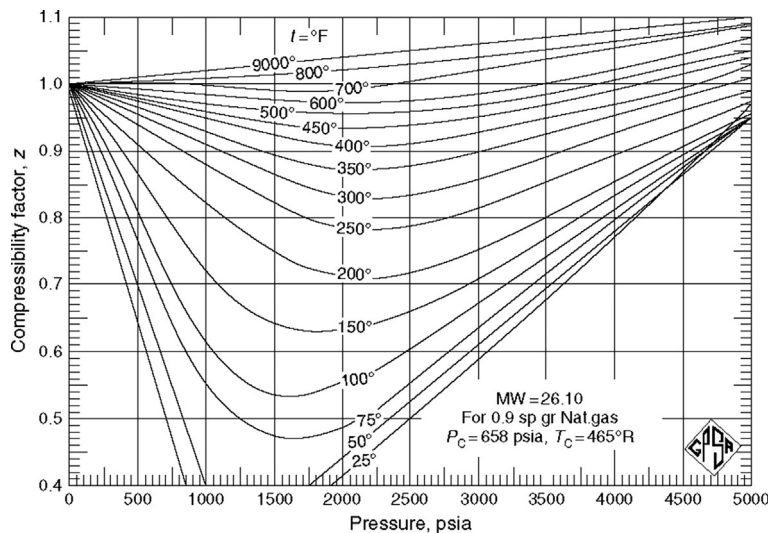


Figure 2.6 Compressibility factor for specific gravity. 0.9 gases. *Courtesy of GPSA Engineering Data Book.*

2.3.5 Liquid Density and Specific Gravity

The specific gravity of a liquid is the ratio of the density of the liquid at 60 °F to the density of pure water at 60 °F.

$$SG = \frac{\rho_1}{\rho_w} \quad (2.10)$$

Where SG = specific gravity of liquid; ρ_1 = density of liquid; ρ_w = density of water at 60 °F.

The density of crude oil is sometimes shown in American Petroleum Institute (API). This term is defined by the equation

$$SG = \frac{141.5}{131.5 + ^\circ\text{API}} \quad (2.11)$$

or

$$^\circ\text{API} = \frac{141.5}{SG} - 131.5 \quad (2.12)$$

In most calculations, the specific gravity of liquids is typically referenced to actual temperature and pressure conditions. [Figure 2.7](#) can be used to approximate how the specific gravity of a liquid decreases with increasing temperature, assuming no phase changes. In most practical pressure drop calculations associated with production facilities, the difference in specific gravity caused by pressure changes is not severe enough to be considered if there are no phase changes.

For hydrocarbons, which undergo significant phase changes, [Figure 2.8](#) can be used as an approximation of the specific gravity at a given pressure and temperature, once the API gravity of the liquid is known.

It should be pointed out that both [Figures 2.7](#) and [2.8](#) are approximations only for the liquid component. Where precise calculation is required for a hydrocarbon, it is necessary to consider the gas that is liberated with decreasing pressure and increasing temperature. Thus, if a hydrocarbon is heated at constant pressure, its specific gravity will increase as the lighter hydrocarbons are liberated. The change in the molecular makeup of the fluid is calculated by a “flash calculation,” which is described in more detail later in this chapter.

2.3.6 Liquid Volume

By definition, 1 API barrel = 42 US gallons at 60 °F

$$\begin{aligned} 1 \text{ API bbl} &= 42 \text{ US gallons} = 35 \text{ UK (Imperial) gallons} = 5.61 \text{ ft}^3 \\ &= 0.159 \text{ m}^3 = 159 \text{ l} \end{aligned}$$

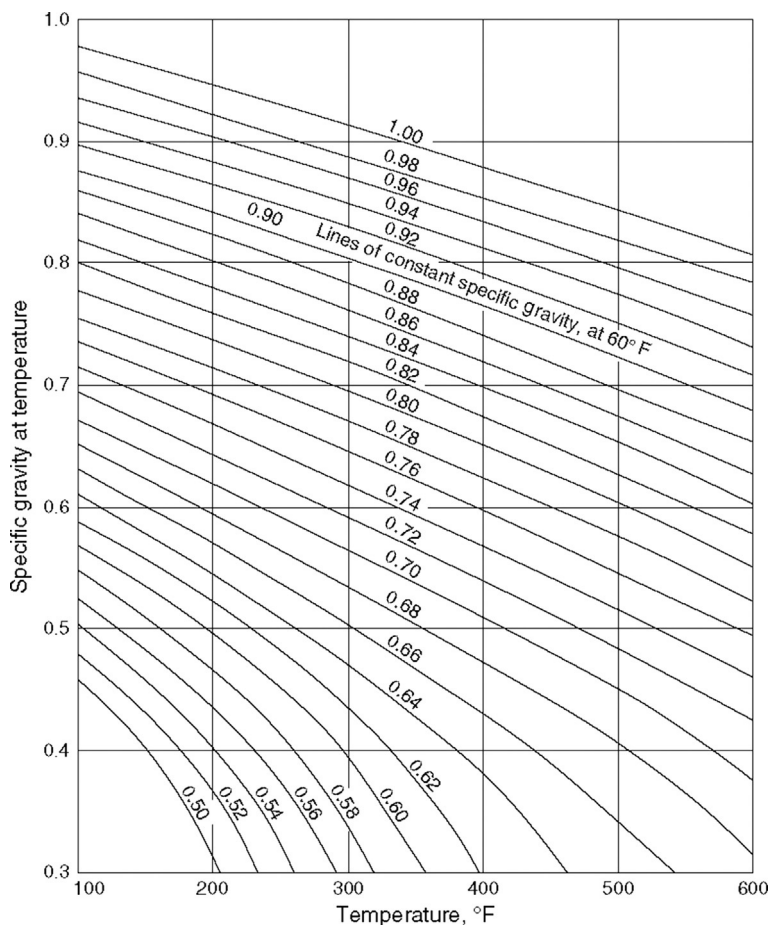


Figure 2.7 Approximate specific gravity of petroleum fractions. *Courtesy of GPSA Engineering Data Book.*

2.3.7 Viscosity

This property of a fluid indicates its resistance to flow. It is an important property used in flow equations and sizing of process equipment. It is a dynamic property in that it can be measured only when the fluid is in motion. Viscosity is a number that represents the drag forces caused by the attractive forces in adjacent fluid layers. It might be considered as the internal friction between molecules, separate from that between the fluid and the pipe wall.

$$1 \text{ centiPoise (cP)} = 0.01 \text{ g/cms}$$

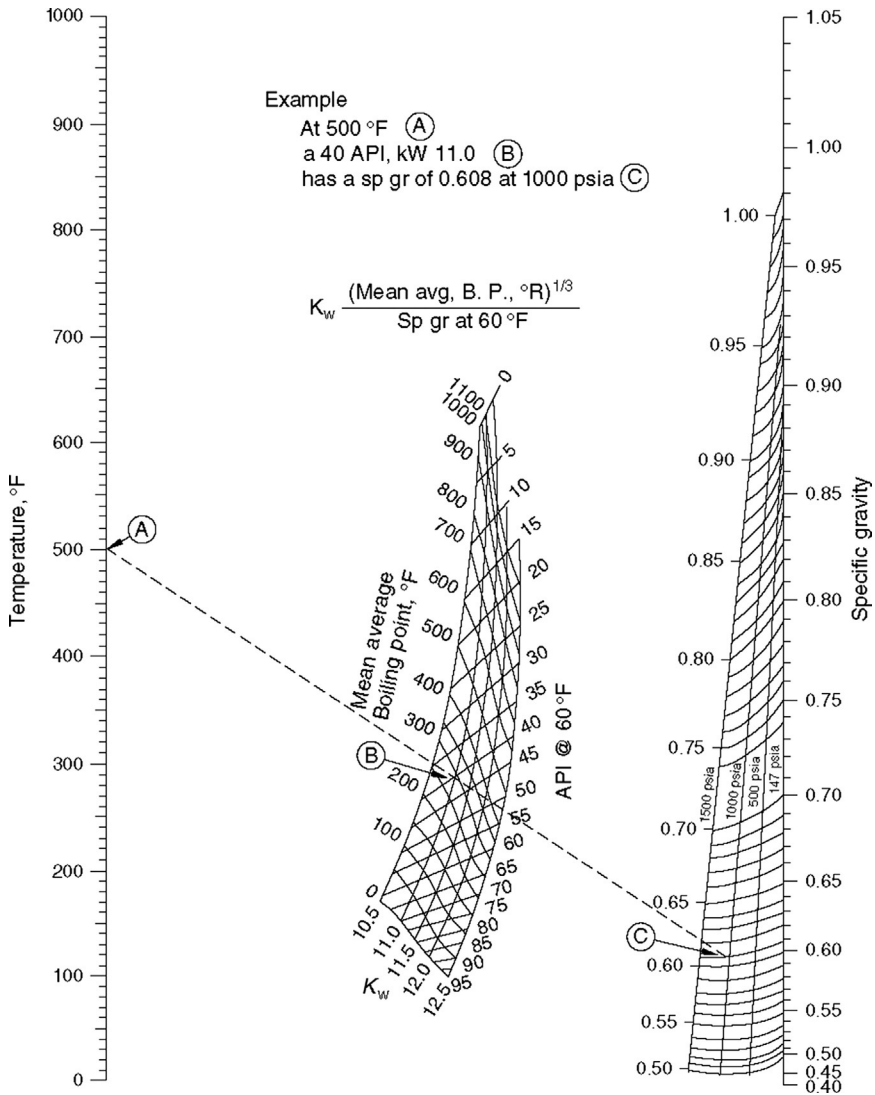


Figure 2.8 Specific gravity of petroleum fractions. *Courtesy of Petroleum Refiner: Ritter, Lenory, and Schweppe, 1958.*

There are two expressions of viscosity: absolute (or dynamic) viscosity, μ , and kinematic viscosity.

These expressions are related by the following equation:

$$Y = \frac{\mu}{\rho} \quad (2.13)$$

where μ = absolute viscosity, centiPoise (cP); Y = kinematic viscosity, centistokes (cSt); ρ = density, g/cm³
and

$$1 \text{ cSt} = 0.01 \text{ cm}^2/\text{s} = 1.0 \times 10^{-6} \text{ m}^2/\text{s}$$

Fluid viscosity changes with temperature. Liquid viscosity decreases with increasing temperature, whereas gas viscosity decreases initially with increasing temperature and then increases with further increasing temperature.

Figure 2.9 can be used to estimate the viscosity of a hydrocarbon gas at various conditions of temperature and pressure if the specific gravity of the gas at standard conditions is known. It is useful when the gas composition is not known; it does not make corrections for H₂S, CO₂, and N₂; and it is useful for determining viscosities at high pressure. Unfortunately, it is an approximate correlation and thus yields less accurate results than other correlations, but for most engineering calculations Figure 2.9 yields results within acceptable limits. When compared to liquid viscosity, gas viscosity is very low due to the relatively large distances between molecules in a gas.

The best way to determine the viscosity of a crude oil at any temperature is by measurement. If the viscosity is not known, Figure 2.10 can be used as a rough approximation. If the viscosity is known at only one temperature, Figure 2.10 can be used to determine the viscosity at another temperature by striking a line parallel to the lines shown. Care must be taken to ensure that the crude does not have its pour point within the temperature range of interest. If it does, its temperature-viscosity relationship may be as shown for crude “B” in Figure 2.11.

Solid phase high-molecular-weight hydrocarbons, otherwise known as paraffins, can dramatically affect the viscosity of the crude sample. The cloud point is the temperature at which paraffins first become visible in a crude sample. The effect of the cloud point on the temperature-viscosity curve is shown for crude “B” in Figure 2.11. This change in the temperature-viscosity relationship can lead to significant errors in estimation. Therefore, care should be taken when estimating viscosities near the cloud point.

The pour point is the temperature at which the crude oil becomes a solid and ceases to flow, as measured by a specific American Society of Testing Materials (ASTM) procedure (D97). Estimations of viscosity near the pour point are highly unreliable and should be considered accordingly.

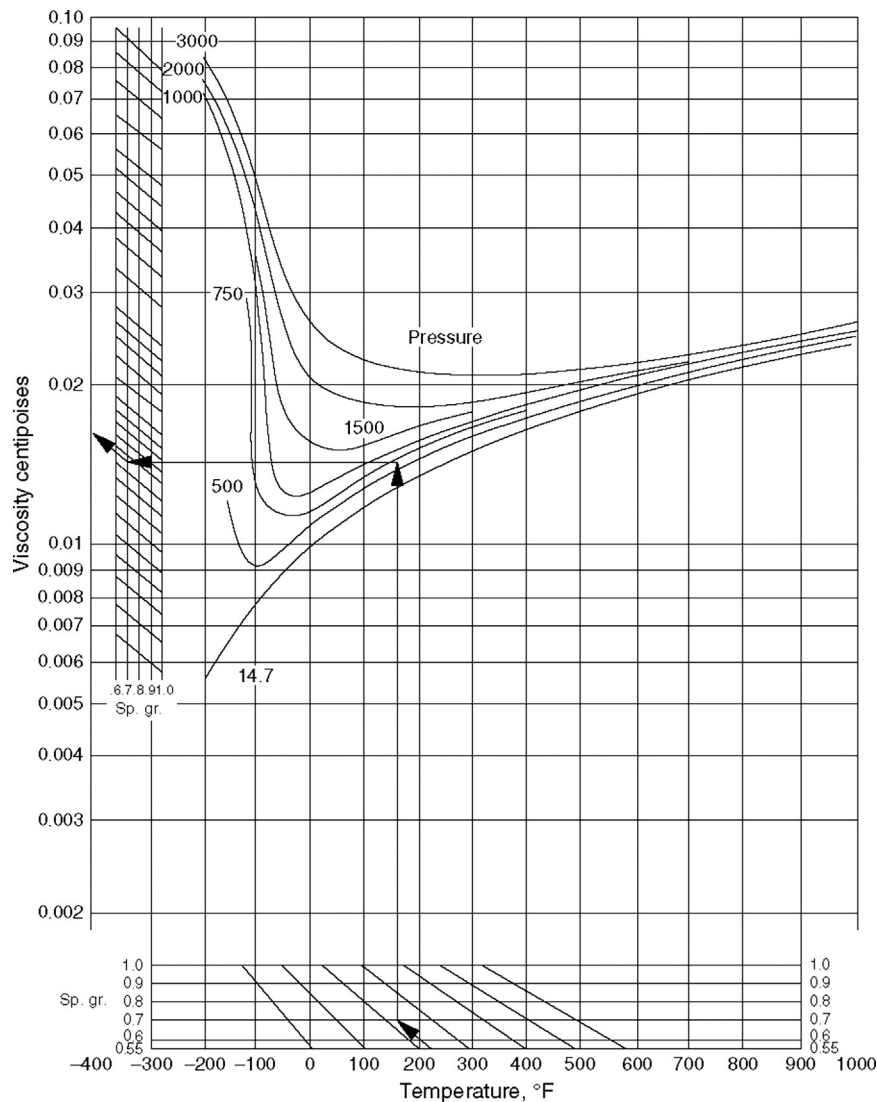


Figure 2.9 Hydrocarbon gas viscosity. *Courtesy of GPSA Engineering Data Book.*

The viscosity of produced water depends on the amount of dissolved solids in the water as well as the temperature, but for most practical situations it varies from 1.5 to 2 cP at 50 °F, 0.7 to 1 cP at 100 °F, and 0.4 to 0.6 cP at 150 °F.

When an emulsion of oil and water is formed, the viscosity of the mixture may be substantially higher than either the viscosity of the oil or that of

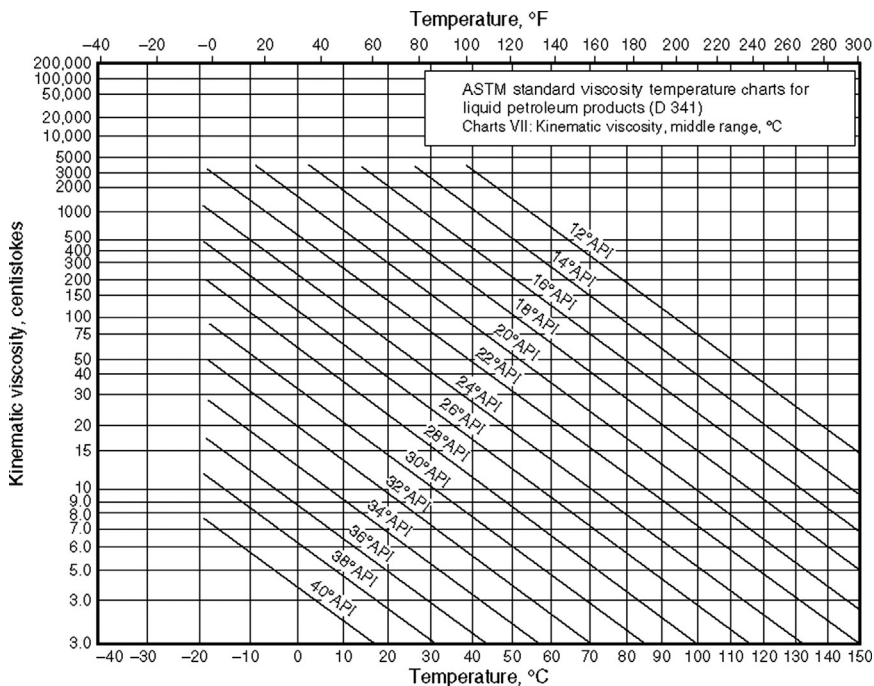


Figure 2.10 Oil viscosity versus gravity and temperature. *Courtesy of Paragon Engineering Services, Inc.*

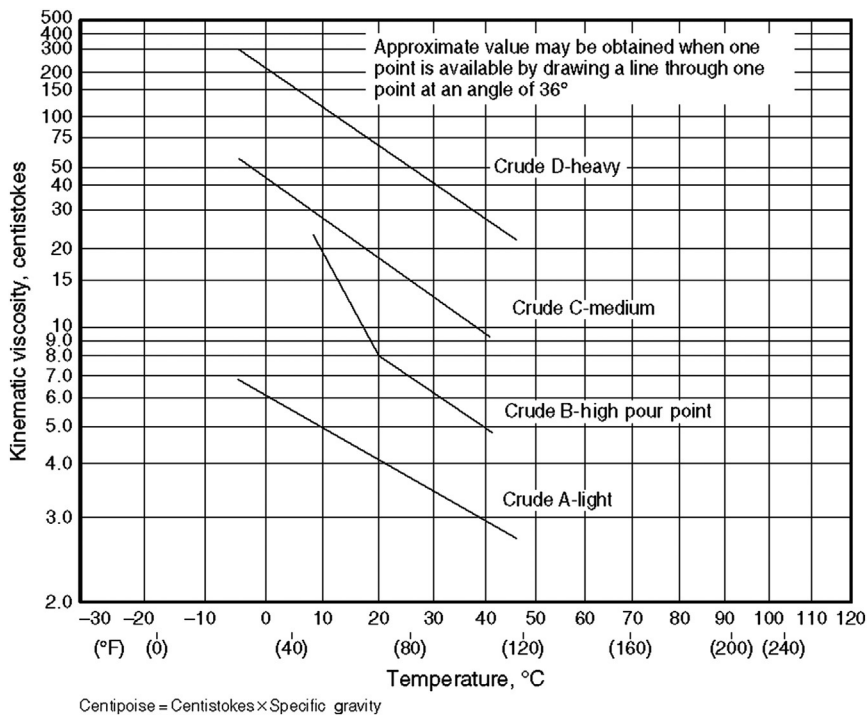


Figure 2.11 Typical viscosity-temperature curves for crude oils. *Courtesy of ASTM D-341.*

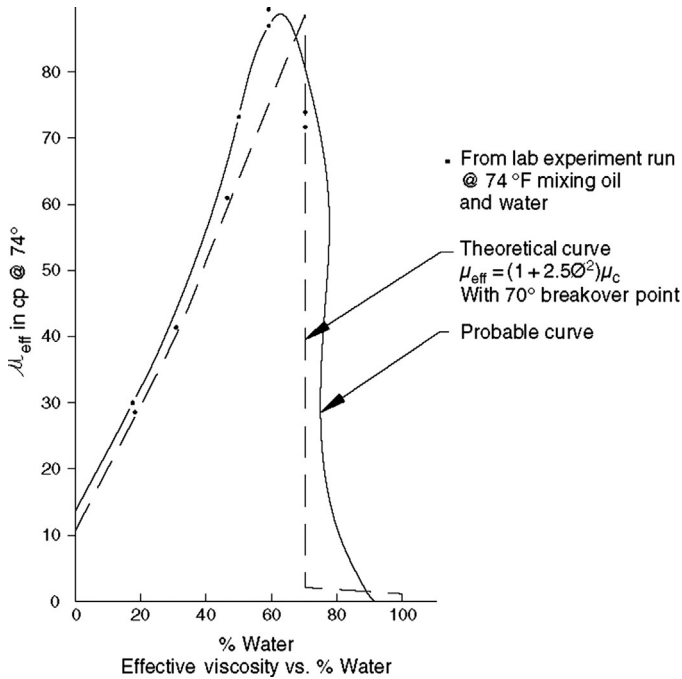


Figure 2.12 Effective viscosity of an example oil/water mixture.

the water. Figure 2.12 shows some experimental data for a mixture of produced oil and water taken from a south Louisiana field. Produced oil and water were mixed vigorously by hand, and viscosity was measured for various percentages of water. For 70% water cut, the emulsion began to break before viscosity readings could be made, and for water cuts greater than this, the oil and water began to separate as soon as the mixing stopped. Thus, at ~70% water cut, it appears that oil ceases to be the continuous phase and water becomes continuous.

The laboratory data plotted in Figure 2.12 agree closely with the modified Vand's equation assuming a 70% breakover point. This equation allows one to determine the effective viscosity of an oil-water mixture and is written in the form

$$\mu_{\text{eff}} = (1 + 2.5\phi + 10\phi^2)\mu_c \quad (2.14)$$

where μ_{eff} = effective viscosity; μ_c = viscosity of the continuous phase; ϕ = volume fraction of the discontinuous phase.



2.4 FLASH CALCULATIONS

2.4.1 Determine Gas and Liquid Compositions

The amount of hydrocarbon fluid that exists in the gaseous phase or the liquid phase at any point in the process is determined by a flash calculation. As explained in Chapter 2, Process Selection (Volume 1), for a given pressure and temperature, each component in the gas phase will depend not only on pressure and temperature, but also on the partial pressure of the component. Therefore, the amount of gas depends on the total composition of the fluids, as the mole fraction of any one component in the gas phase is the function of the mole fraction of every other component in this phase.

This is best understood by assigning an equilibrium “ K ” value to each component. The K value is a strong function of temperature and pressure and of the composition of the vapor and liquid phase. It is defined as

$$K_N = \frac{V_N/V}{L_N/L} \quad (2.15)$$

where K_N = constant for component N at a given temperature and pressure; V_N = moles of component N in the vapor phase; V = total moles in the vapor phase; L_N = moles of component N in the liquid phase; L = total moles in the liquid phase.

The Gas Processors Suppliers Association (GPSA) present graphs of K values for the important components in a hydrocarbon mixture such as that shown in Figure 2.13. The K values are for specific “convergence” pressure. A procedure in the GPSA’s Engineering Data Book for calculating convergence pressure is based on simulating a binary fluid system with the lightest hydrocarbon component, which makes up at least 0.1 mol% in the liquids and a pseudoheavy component having the same average weight and critical temperature as the remaining heavier hydrocarbons. The convergence pressure is then read from a graph of convergence pressure versus operating temperature for common pseudobinaries. In most oil-field applications the convergence pressure is between 2000 and 3000 psia, except at very low pressures, where a psia between 500 and 1500 is possible. If the operating pressure is much less than the convergence pressure, the equilibrium constant is not greatly affected by the choice of convergence pressure. Therefore, using a convergence pressure of 2000 psia is a good first approximation for most flash calculations. Where greater precision is required, the

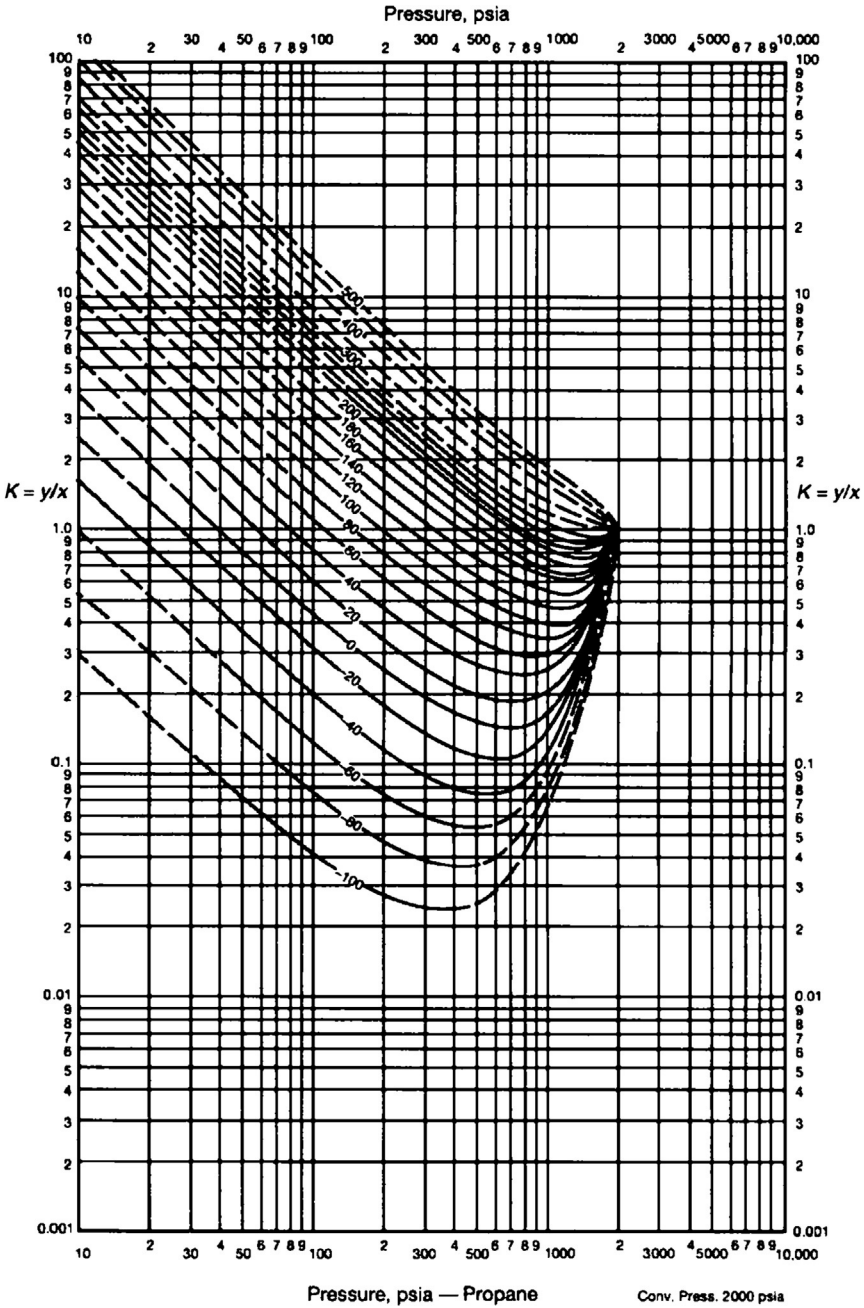


Figure 2.13 "K" values for propane. Courtesy of GPSA Engineering Data Book.

convergence pressure should be calculated. If K_N for each component and the ratio of total moles of vapor to total moles of liquid (V/L) are known, then the moles of the component N in vapor phase (V_N) and the moles in the liquid phase (L_N) can be calculated from

$$V_N = \frac{K_N F_N}{\frac{1}{V/L} + K_N} \quad (2.16)$$

$$L_N = \frac{F_N}{K_N(V/L) + 1} \quad (2.17)$$

where F_N = total moles of component N in the fluid.

To solve either Equations (2.16) or (2.17), it is necessary to first know the quantity (V/L), but since both V and L are determined by summing V_N and L_N , it is necessary to use an iterative solution. This is done by estimating (V/L), calculating V_N and L_N for each component, summing up to obtain the total moles of gas (V) and liquid (L), and then comparing the calculated (V/L) to assumed value.



2.5 CHARACTERIZING THE FLOW STREAM

Once a flash calculation is made and the molecular composition of the liquid and gas components have been determined, it is possible to determine the properties and flow rates of both the gas and the liquid streams.

The molecular weight of a stream is calculated from the weighted average gas molecular weight given by

$$\text{MW} = \sum [V_N \times (\text{MW})_N] \quad (2.18)$$

The gas's specific gravity can be determined from the molecular weight from Equation (2.7). If the flow of the inlet stream is known in moles per day, then the number of moles per day of gas flow can be determined from

$$V = \frac{F}{1 + \frac{1}{V/L}} \quad (2.19)$$

where V = gas flow rate, mol/day; F = total stream flow rate, mol/day; L = liquid flow rate, mol/day.

Once the mole flow rate of gas is known, then the flow rate in standard cubic feet can be determined by recalling that one mole of gas occupies 380 ft³ at standard conditions. Therefore,

$$Q_g = \frac{380 V}{1,000,000} \quad (2.20)$$

where Q_g = gas flow rate, MMscfd.

The molecular weight of the liquid stream is calculated from the weighted average liquid component molecular weight given by

$$MW = \frac{\sum [L_N \times (MW)_N]}{L} \quad (2.21)$$

Remembering that the weight of each component is the number of moles of that component times its molecular weight, the specific gravity of the liquid is given by

$$SG = \frac{\sum [L_N \times (MW)_N]}{\sum \frac{[L_N \times (MW)]_N}{(SG)_N}} \quad (2.22)$$

The liquid flow rate in barrels per day can be derived from

$$Q_1 = \frac{L \times (MW)}{350(SG)} \quad (2.23)$$

where Q_1 = liquid flow rate, bpd; SG = specific gravity of liquid (water = 1).

Often, the designer is given the mole fraction of each component in the feed stream but is not given the mole flow rate for the stream. It may be necessary to estimate the total number of moles in the feed stream (F) from an expected stock-tank oil flow rate. As a first approximation, it can be assumed that all the oil in the stock tank can be characterized by the C_7^+ component of the stream. Thus, the feed rate in moles per day can be approximated as

$$L \cong \frac{350(SG_7)Q_1}{(MW)_7} \quad (2.24)$$

where L = liquid flow rate, mol/day; $(SG)_7$ = specific gravity of C_7^+ ; $(MW)_7$ = molecular weight of C_7 ; Q_1 = flow rate of liquid, bpd.

The mole flow rate of the feed stream is then calculated as

$$F = \frac{L}{(\text{mole fraction})_7} \quad (2.25)$$

where F = flow rate feed stream, mol/day; (mole fraction)₇ = mole fraction of the C_7^+ component in the feed stream.

The flash calculation could then proceed. The calculated flow rates for each stream in the process could then be used in a ratio to reflect the error between assumed stock-tank flow rate and desired stock-tank flow rate. Refer to Surface Production Operations, Volume 1, for a complete example using this hand calculation method.



2.6 USE OF COMPUTER PROGRAMS FOR FLASH CALCULATIONS

The iterative manual flash calculation detailed in the previous sections shows one of many methods for calculating equilibrium conditions. Flash calculations are inherently rigorous and best performed by sophisticated simulation software, such as HYSIM or similar programs.



2.7 APPROXIMATE FLASH CALCULATIONS

Sometimes it is necessary to get a quick estimate of the volume of gas that is expected to be flashed from an oil stream at various pressures.

Figure 2.14 was developed by flashing several crude oils of different gravities at different pressure ranges. The curves are approximate. The actual shape would depend on the initial separation pressure, the number and pressure of intermediate flashes, and the temperature.

Use of the curve is best explained by an example. Suppose a 30° API crude with a Gas-Oil-Ratio (GOR) of 500 is flashed at 1000, 500, and 50 psia before going to a stock tank. Roughly 50% of the gas that will eventually be flashed from the crude, or 250 ft³/B, will be liberated as gas in the 1000-psia separator. Another 25% (75–50%), or 125 ft³/B, will be separated at 500 psia, and 23% (98–75%), or 115 ft³/B, will be separated at 50 psia. The remaining 10 ft³/B (100–98%) will be vented from the stock tank.

It must be stressed that Figure 2.14 is only to be used where a quick approximation, which could be subject to error, is acceptable. It cannot be used for estimating gas flashed from condensate produced in gas wells.



2.8 OTHER PROPERTIES

Once the equilibrium conditions (and, therefore, the gas and the liquid compositions) are known, several very useful physical properties

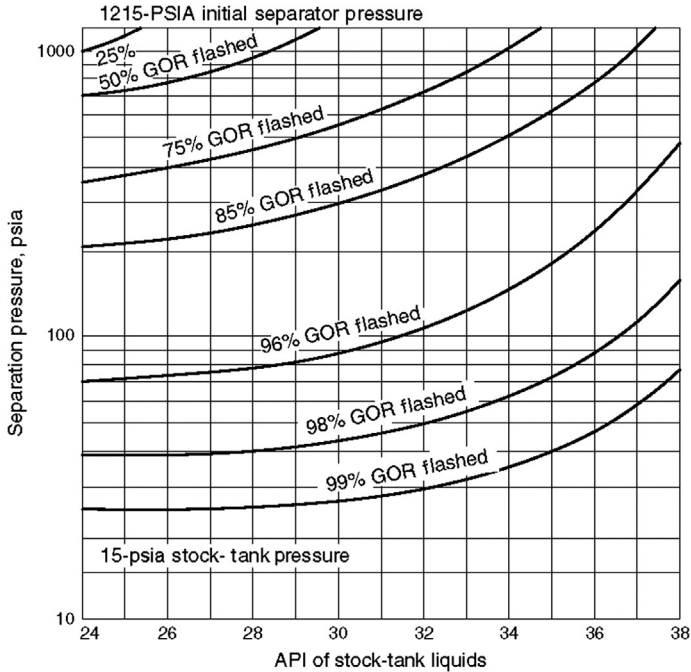


Figure 2.14 Preliminary estimation of % GOR flashed for given API of stock-tank liquids and separation pressures—Gulf Coast Crudes.

are obtainable, such as the dew point, the bubble point, the heating value (net and gross), and k , the ratio of gas-specific heats. These properties are described next:

2.8.1 Dew Point: the point at which liquid first appears within a gas sample.

A more precise definition of the dew point makes a distinction between the hydrocarbon dew point, which represents the condensation of a hydrocarbon liquid, and the water dew point, which represents the condensation of liquid water. Often, sales gas contracts specify control of the water dew point for hydrate and corrosion control and not the hydrocarbon dew point. In such cases, hydrocarbons will often condense in the pipeline as the gas cools (assuming that separation has occurred at a higher temperature than ambient), and provisions to separate this condensate must be provided.

2.8.2 Bubble Point: the point at which gas first appears within a liquid sample.

2.8.3 Net Heating Value: heat released by combustion of gas sample with water vapor as a combustion product; also known as the lower heating value.

2.8.4 Gross Heating Value: heat released by combusting of gas sample with liquid water as a combustion product; also known as the higher heating value.

2.8.5 k: ratio of heat capacity at constant pressure (CP) to heat capacity at constant volume (CV). Often, this is used in compressor calculation of horsepower requirement and volumetric efficiencies. This ratio is relatively constant for natural gas molecular weight and ranges between 1.2 and 1.3 (see [Figure 2.15](#)).

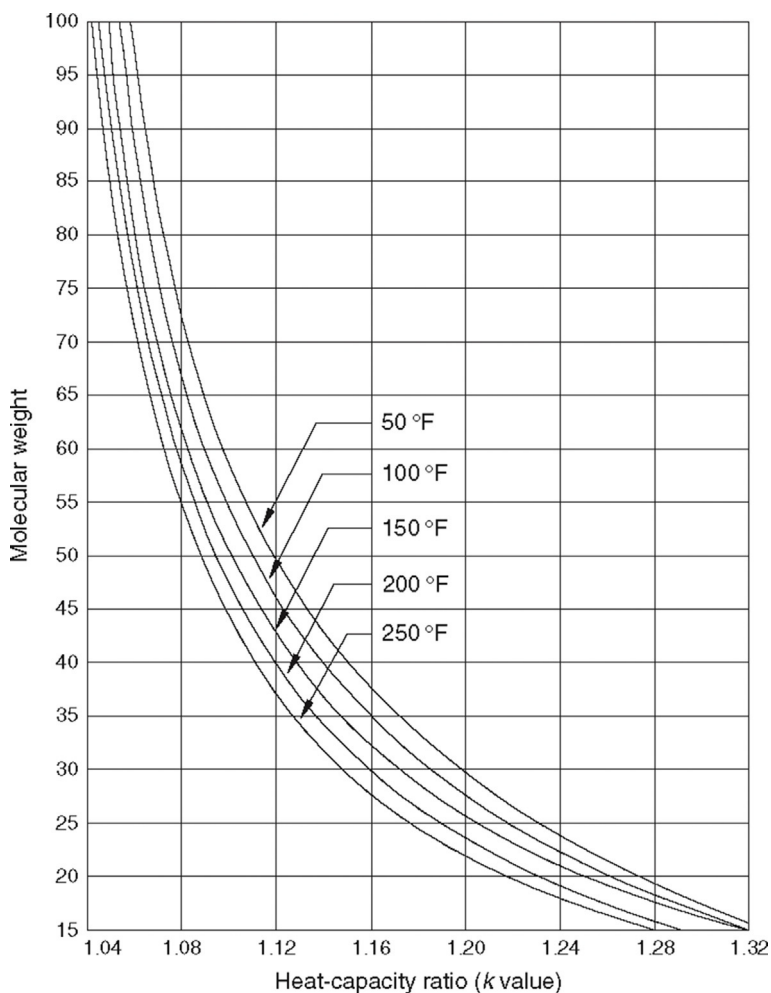


Figure 2.15 Approximate heat-capacity ratios of hydrocarbon gases. *Courtesy of GPSA Engineering Data Book.*

2.8.6 Reid Vapor Pressure: the bubble point can also be referred to as the “true vapor pressure.” A critical distinction lies here between the true vapor pressure and the Reid vapor pressure (RVP). The RVP is measured according to a specific ASTM standard (D323) and lies below the true vapor pressure. The approximate relationship between the two pressures is shown in [Figure 2.16](#). (Note that an RVP below atmospheric pressure does not indicate that vapors will be absent from a sample at atmospheric pressure.)

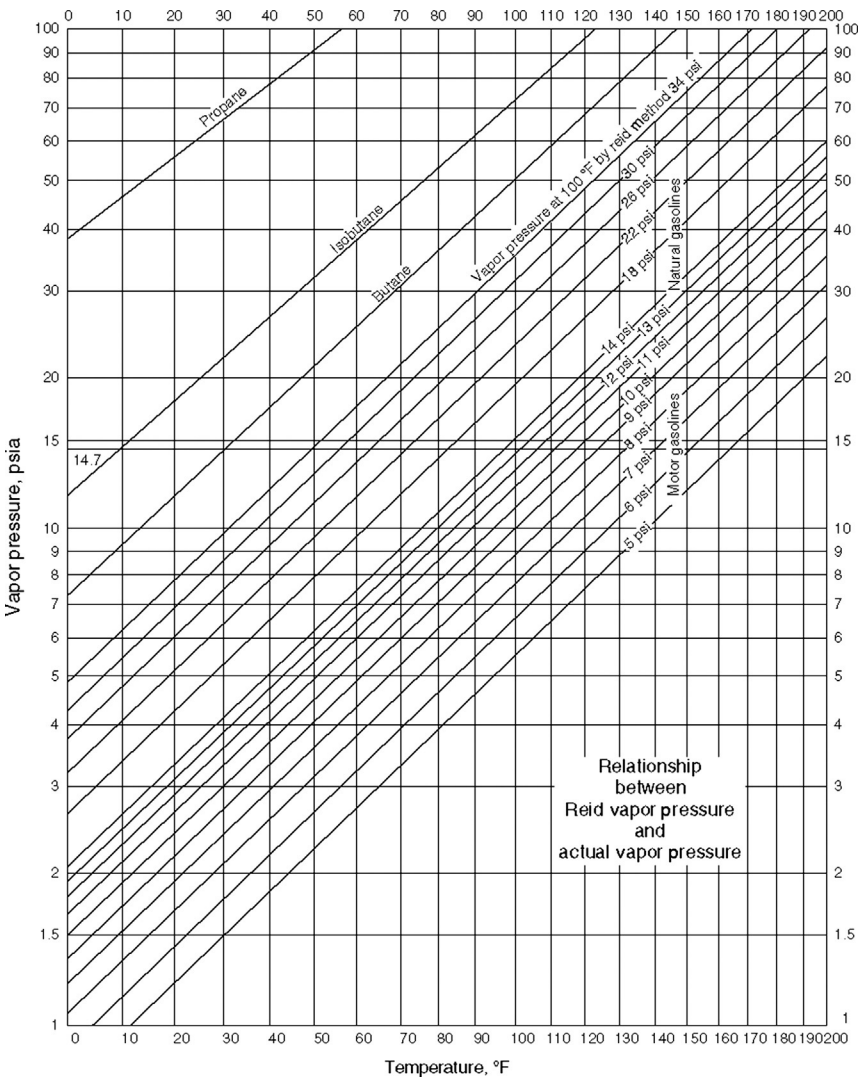


Figure 2.16 Relationship between Reid vapor pressure and actual vapor pressure. Courtesy of GPSA Engineering Data Book.

2.9 PHASE EQUILIBRIUM

A basic representation of the equilibrium information for a specific fluid composition can be found in a P - H (pressure–enthalpy) diagram, which is highly dependent on the sample composition. This diagram can be used to investigate thermodynamic fluid properties as well as their thermodynamic phenomena, such as retrograde condensation and the Joule–Thomson effect. Please note, however, that a P - H diagram is likely to be available only for a single component of the mixture unless the diagram is created by simulation software packages such as those mentioned above. A P - H diagram for propane is shown in Figure 2.17; a P - H diagram for a 0.6 specific gravity natural gas is shown in Figure 2.18.

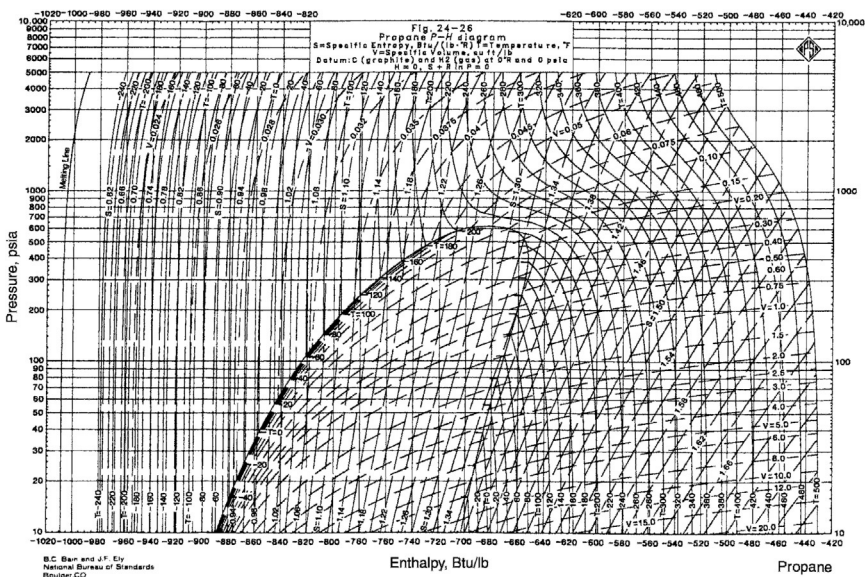


Figure 2.17 A P - H diagram for propane. Courtesy of GPSA Engineering Data Book.

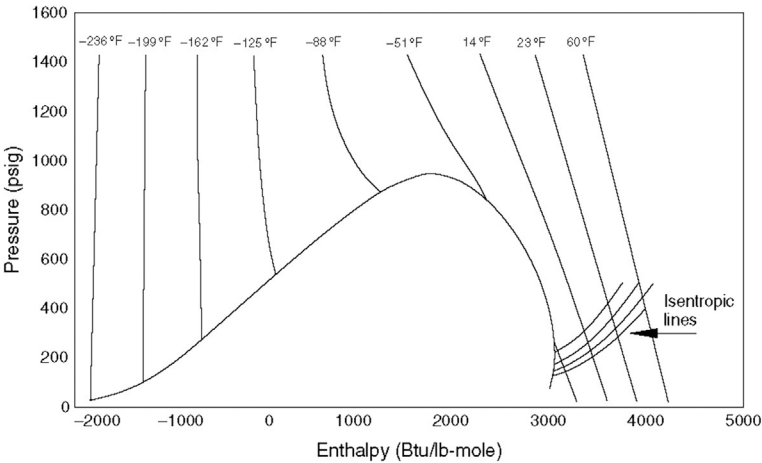


Figure 2.18 A *P-H* diagram for 0.6 specific gravity natural gas.



Heat Transfer Theory



3.1 OBJECTIVES

The objective of this chapter is to help the reader understand the role of heat exchangers (HEX) in upstream operations. In addition, this chapter allows the reader to

- Learn what information is required to
 - define the service and range of operation
 - develop stream properties
- Evaluate heat exchanger operation and performance
- Understand principles in heat exchanger design
- Explore basic mechanical standards covering HEX design, including
 - American Society of Mechanical Engineers (ASME) Section VIII, Division I—for most exchangers
 - Tubular Exchanger Manufacturers Association (TEMA) Class “R”—for petroleum processing facilities
 - American Petroleum Institute (API) Standards 660, 661, 662, and 632
- Determine what TEMA type is appropriate for each type of service
- Learn how to handle known design/modification challenges
- Know what information is required for heat exchanger design and evaluation
- Explore basic maintenance practices for heat exchangers

This chapter concentrates on basic heat transfer theory and process heat duty. [Chapter 4](#) discusses heat exchanger nomenclature and heat exchanger types and configurations such as:

- Shell-and-tube
- Double pipe
- Multitube hairpin
- Plate-and-frame
- Plate-fin (brazed aluminum)
- Indirect-fired heaters

- Direct-fired heaters
- Waste heat recovery
- Air-cooled exchangers
- Cooling towers



3.2 WHAT IS A HEAT EXCHANGER?

A heat exchanger is a device built for efficient heat transfer from one medium to another, whether the media are separated by a wall so that they never mix, or they are in direct contact.

Heat exchangers are widely used in both upstream and downstream facilities. They may be classified according to their flow arrangement. For example:

- Co-current flow (parallel flow)
Two fluids enter the exchanger at the same end and travel parallel to each other to the other side.
- Counter-current flow (counter-flow)
Two fluids enter the exchanger from opposite ends.

3.2.1 Commonly Used Types of Heat Exchangers

Shell-and-Tube

A shell-and-tube heat exchanger consists of a bundle of tubes within a shell. One fluid that needs to be heated or cooled flows through the tubes, and the second fluid runs over the tubes to provide the heat or absorb the heat required.

Plate-and-Frame

A plate-and-frame exchanger is composed of multiple, thin, slightly separated plates that have very large surface areas relative to their size for heat transfer.

Plate-Fin

A plate-fin exchanger uses fins to increase the efficiency of the unit. Designs include cross-flow and counter-flow coupled with various fin configurations (straight, offset, or wavy).

Air-Cooled

An air-cooled heat exchanger consists of a bundle of tubes with air passing around the outside.

3.2.2 Heat Exchangers—The Bad News

The leading causes of lost production opportunities (LPOs) are shown in [Figure 3.1](#). Heat exchangers are susceptible to problems such as fouling, corrosion, and leaks despite having no moving parts. These problems can also limit unit throughput.

3.2.3 General Considerations

The starting points for heat exchanger design are:

- (1) ASME Boiler and Pressure Vessel Codes:
 - Section VIII Division 1 for most exchangers
 - Section I for “fired” steam generators
- (2) Standards of the Tubular Exchanger Manufacturers Association (TEMA):
 - Class “R” for severe petroleum and related processing applications
 - Class “C” for commercial and general process applications
 - Class “B” for chemical process service
- (3) API standards:
 - API 660 Shell-and-Tube Exchangers for General Refinery Services
 - API 661 Air-Cooled Heat Exchangers for General Refinery Services
 - API 632 Winterization of Air-Cooled Heat Exchangers
 - API 662 Plate Heat Exchangers for General Refinery Services

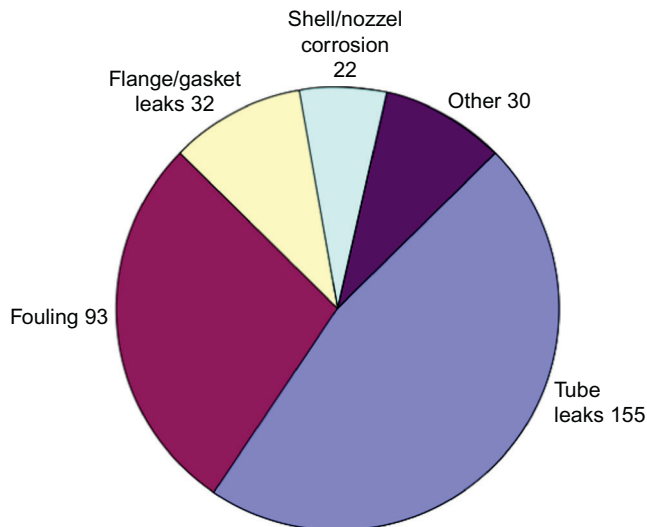


Figure 3.1 Typical causes of heat exchanger LPOs (based on 332 cases in a 5-year study; courtesy of Chevron USA).

None of the above standards address temperature differences and thermal stresses. ASME Section VIII, Division 1 covers mechanical design of pressure-containing parts (e.g., shells and channel) and some aspects of large flanges. Section VIII ignores thermal stresses and deformation, which is of consequence to flange tightness more than about 250 °F, or 121 °C, and tube sheet integrity at very high temperatures. ASME Section I provides rules for the design of stayed tube sheets, which are necessary to keep tube sheets thin and cooled by water quenching in the presence of combustion or other hot gases.

TEMA addresses shell-and-tube exchangers only and covers nomenclature, fabrication tolerances, standard clearances, minimum plate thickness, and tube sheet design rules.

API standards reference the aforementioned standards and several others relating to piping, welding materials, and nondestructive examination; require many design features that are universally applicable (e.g., seal bars); and provide a checklist of decisions that must be made by the purchaser.

Most companies use API 660 for shell-and-tube exchangers and API 661 for air-cooled exchangers. There is little demand and incentive to use plate exchangers in downstream refinery service, but they are widely used in upstream operations, especially in offshore installations where a small footprint is required.

Industry standards are limited to moderate pressures and temperatures and do not cover subjects that are beyond the control of manufacturers, such as fouling, corrosion, vibration, leak tightness of flanges, and tube ruptures.

3.2.4 Company Engineering Standards and Specifications

Company engineering standards and specifications address subjects that are applicable to most exchangers and that are beyond the scope of industry standards. They typically reference and supplement API 660 for shell-and-tube exchangers and address additional requirements concerning fouling, corrosion, tube vibration and tube rupture, and leak tightness of flanges. In addition, they cover exchangers using special materials or thick-wall construction. Many company standards reference and supplement API 661 for air coolers. Company standards also cover additional topics, such as double pipe, hairpin, and plate-and-frame exchangers. The standards and specifications ensure optimum design and eliminate guesswork.

3.2.5 Tube Vibration and Tube Rupture

Prevention of tube vibration and tube rupture must be integrated with layout, process design, and mechanical design concerns. TEMA addresses flow-induced vibration in detail. Figure 3.2 shows tube vibration mechanisms, and Figure 3.3 illustrates where tube vibrations may occur.

The 2001 edition of the ASME Code incorporates ASME Code 1998 with the 1999 addendum allowing for maximum allowable stresses based on a factor of 3.5 instead of the traditional design factor of 4.0. Prior to the 1999 addendum, all pressure vessels and process piping was required to be hydrostatic tested to 150% of stated design pressure. For example, if a pressure vessel's design pressure is 100 psig, then the hydrostatic test pressure must be 150 psig; so $100/150 = 2/3$. This is known as the “2/3 rule.” API RP 521 states that if the design pressure of the low-pressure side of a heat exchanger is within 2/3 of the design pressure of the high pressure side, then one would not have to include tube rupture as a credible relieving scenario. API's reasoning is a catastrophic failure of a tube would not occur if the two design pressures were so close. API extended this “rule” by specifying that the entire system be within the “2/3 rule,” not only the shell-and-tube of the exchanger. That is, if there is a vessel after the exchanger tied

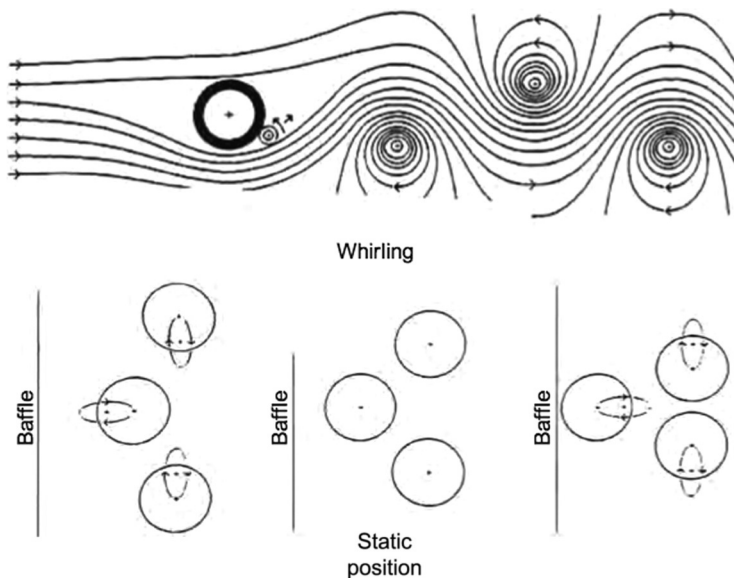


Figure 3.2 Tube vibration mechanisms.

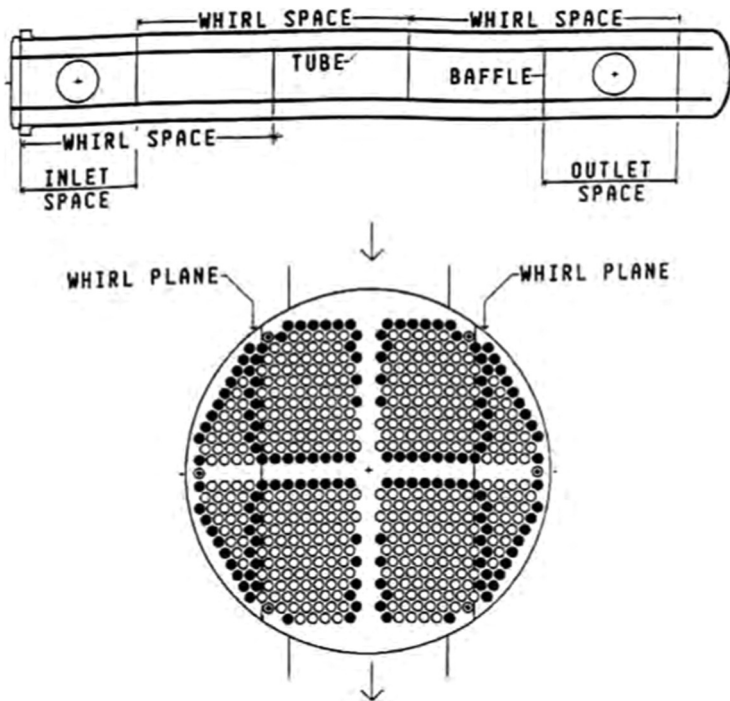


Figure 3.3 Where tube vibrations may occur.

directly into the low-pressure side of the exchanger, then its design pressure must also be within 2/3 of the high-pressure side of the exchanger.

After the 1999 addendum, the ASME Code required the hydrostatic test pressure of pressure vessels to be 130% of the stated design pressure. For example, if a pressure vessel's design pressure is 100 psig, then the hydrostatic test pressure must be 130 psig; so $100/130 = "10/13 \text{ rule}."$ Therefore, to avoid a tube rupture as a credible scenario, the design pressure of the low-pressure side has to be within 10/13 of the design pressure of the high-pressure side, including the system. Again, the test pressure of **equipment** is allowed to be tested at 130% of the stated design pressure, but process **piping** must still be tested at 150%. The ASME Code does not prohibit hydrostatic testing of equipment at 150% of the stated design pressure. Therefore, if one tests at 150%, then one can still use the "2/3 rule" to determine if tube rupture is a credible scenario.

Most companies' design standards improve on the "2/3 rule." Increasing the low-pressure side design pressure will have little impact on the likelihood

of a tube rupture. However, by raising the low-pressure side design pressure to a point where the low-pressure side test pressure is equivalent to the high-pressure side design pressure ensures that damage to the exchanger due to a tube rupture is extremely unlikely.

Design standards have evolved over the years due to investigation of catastrophic plant incidents and large-scale laboratory tests by Southwest Research Institute 1977–1979.

Tube rupture industry standards and recommended practices include the following:

- ASME Section VIII Division 1 UG-133 (d) Code case VIII-80-56 (6/25/80) requires relief provisions for steam boilers only.
- API Recommended Practice 521 discusses inadequacy of ASME code rule and is not helpful. It recommends considering complete break of one tube or break at one end, which would be the worst-case scenario.
- Standards recommend treating high-pressure, two-phase fluid as all gas (liquid is moving slower). Refer to [Figure 3.4](#).

High-pressure closures are not commodity items and must be custom designed. These closures are not easy to design. Using a manufacturer with a good track record in the design of these specialty items is recommended. Many company standards recommend the use of a high-pressure closure when the following conditions occur:

Field Units

Pressure (psi) \times diameter (in.) is $>75,000$

SI Units

Pressure (bar) \times diameter (mm) is $>131,345$

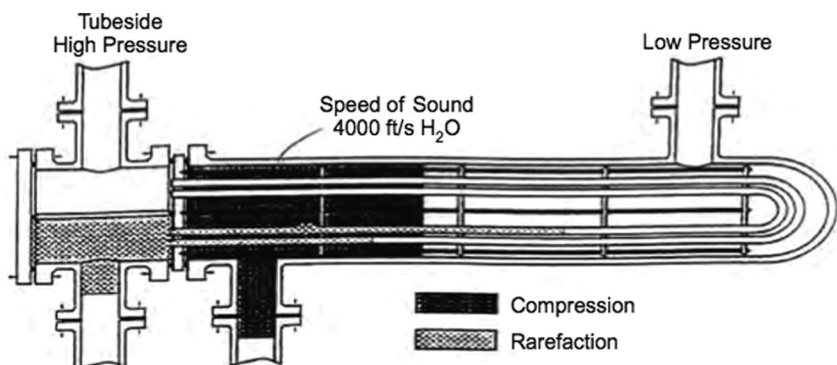


Figure 3.4 Tube rupture transient schematic.

3.2.6 Acoustic Resonance and Vibration

Tube banks located in shell-and-tube process heat exchangers and exposed to cross-flow have a low resistance to acoustic resonance and vibration. Such resistance can be increased substantially by placing acoustic baffles inside the tube bank. The baffles distort the acoustic waves and thereby increase the resistance of the tube bank to acoustic resonance and vibration. The locations and number of baffles play an important role in the degree of their effectiveness in suppressing acoustic waves. The effects of the baffles raising the vibratory thresholds, or reducing the noise levels at resonance, have been evaluated. It has been shown that one or two baffles placed inside a tube bank raises the threshold to acoustic vibration significantly. For example, one baffle can raise the vibratory threshold of up to a factor of about 4 for acoustic modes 1 to 5, while two baffles primarily affect the higher modes (modes 4 and 5) by raising the threshold values by an additional factor of 3 [1,2]. Optimum baffle locations suppressing acoustic resonance and vibration or reducing noise levels at resonance can be determined. General procedures have been developed that can be used for the evaluation of the effects of acoustic baffles for any desired

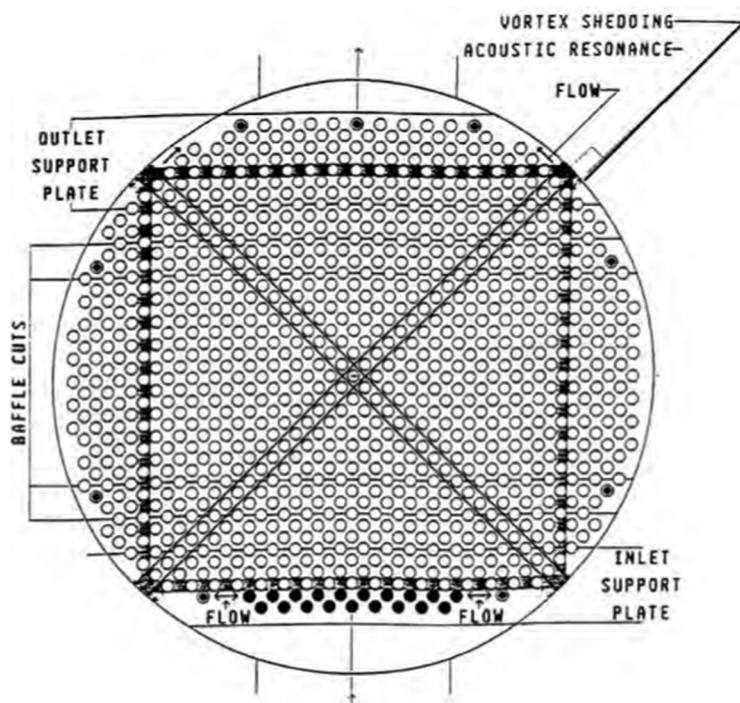


Figure 3.5 Acoustic resonance.

baffle configuration in any flow-channel geometry. Figure 3.5 is an example of a tube bank with proper baffles placement that increases the threshold to acoustic resonance and vibration. Mitigating acoustic resonance and vibration is beyond the scope of this text but should be investigated.



3.3 PROCESS SPECIFICATION

3.3.1 Process Specification Sheet

The process specification sheet describes the process requirements and design preferences for a heat exchanger's design. It is a communication tool between the company and the heat exchanger manufacturer. The process specification usually includes

- A completed API 660 specification sheet or company design data sheet
- A completed API 660 checklist
- Additional items not addressed on either form

3.3.2 Design Data Sheet

The design data sheet is used to

- Define a new heat exchanger service or provide sufficient information to analyze an existing exchanger
- Allow information flow in both directions
 - The company provides information the manufacturer needs
 - The company receives the following from the manufacturer:
 - Mechanical design information, including fabrication drawings
 - Heat transfer and pressure drop performance information
 - Vibration analysis results

3.3.3 Required Physical/Thermal Properties

The required physical/thermal properties are dependent on the phase(s) for each stream.

The designer should use the standard form from API 660 Appendix D and then add additional information as required. Areas of concern include performance requirements and stream data.

For each phase, or for single-phase fluids, the following is required:

- Density lb/ft^3 (kg/m^3)
- Viscosity cp (Pa s)
- Specific heat (capacity) $\text{Btu}/(\text{lb}\cdot^\circ\text{F})$ [$\text{J}/(\text{kg}\cdot^\circ\text{C})$]
- Thermal conductivity $\text{Btu}/(\text{h}\cdot^\circ\text{F ft})$ [$\text{W}/(\text{m}\cdot^\circ\text{C})$]

For boiling fluid, the following is required:

- Vapor molecular weight
- Weight fraction vapor
- Liquid and vapor phase enthalpies
- Critical temperature and pressure
- Inlet thermal conductivity based on inlet temperature $\text{Btu}/(\text{h } ^\circ\text{F ft})$ ($\text{W}/(\text{m } ^\circ\text{C})$)
- Surface tension over the range extending between inlet temperatures to dew point (usually at two pressures spanning the operation)

For condensing fluid, the following is required:

- Vapor molecular weight
- Weight fraction vapor
- Stream enthalpy
- Surface tension (especially for finned tubes) over the range of the two fluids through the exchanger or through the range where the phase exists

Heat transfer analysis uses basic formulas and computer programs. Much of the film coefficient information is proprietary. Pressure drop considerations for single-phase tube-side are straightforward. Every other mode, shell-side or phase change, is difficult and is generally obtained from proprietary sources. An important relationship is one between heat transfer and pressure drop. The designer must also consider system economics. One must allocate the streams to minimize equipment cost and operating cost. Mechanical and materials constraints include:

- Minimizing
 - Initial equipment cost [capital expense (CAPEX)] and
 - Operating cost [operating expense (OPEX)]
 - Thermal expansion
- Flange and tube leaks
- Safety
- Materials versus process conditions

Fouling information is highly proprietary. The designer must understand and control fouling to avoid its high costs.



3.4 PRESSURE DROP CONSIDERATIONS

3.4.1 Key Parameter for Design and Troubleshooting

The allowable pressure drop should be set by process constraints. Key pressure drop considerations include:

- Evaluating design
 - Determining flow path length

- Calculating pressure gradient ($\Delta P/L$)
- Determining what is a reasonable pressure drop, depending on service
- Minimize cost system. Restricted? Why? Impact limiting side
- Impact of fouling
- Evaluating performance
 - Compare to design
 - Another indicator of fouling
 - What is time trend?

Ways to adjust pressure drop:

Tube side

- Number of tubes per pass
- Tube length
- Number of tube passes
- Number of shells and routing

Shell side

- Flow path length (TEMA type, tube length)
- Baffles (style, cut, spacing)
- Number of shells and routing

Other ways

- Install tube inner diameter inserts
- Change tube pitch

For single-phase and two-phase (still predominantly transferring sensible heat) streams, the pressure drop is simply calculated by multiplying the pressure gradient by the appropriate flow path length as shown in the following equation:

$$\Delta P(\text{psi}) = \frac{\Delta P}{L}(L) \quad (3.1)$$

The resulting pressure drop is for the overall exchanger, including nozzles, channels, heads, or U bends. For two-phase streams, the density is the homogeneous density, assuming no liquid holdup.

Typical Pressure gradients ($\Delta P/L$)

For liquids (HC and H_2O):

Tube side: 0.2–0.3 psi/ft (4524–6784 Pa/m)

Shell side: 0.4–0.6 psi/ft (9048–13,570 Pa/m)

For all fluids including two phase and gases:

Tube side: $(0.05\text{--}0.08) \times (\text{density, lb/ft}^3)^{1/3} (1131\text{--}1809) \times [(\text{density, kg/m}^3)^{1/3}]/2.52$ (Pa/m)

Shell side: $(0.1-0.15) \times (\text{density, lb/ft}^3)^{1/3} (2262-1809) \times [(\text{density, kg/m}^3)^{1/3}] / 2.52 \text{ (Pa/m)}$

Flow path lengths, ft or m (L)

Tube side: tube straight length—number of tube passes

Shell side: axial length (for E shell tube straight length)

Pressure drop can be estimated for new process conditions based on the Darcy-Weisbach flow equation. The equation, expressed in feet (meters) of liquid, is expressed as follows:

$$h = \frac{fL}{D} \frac{V^2}{2g} = \left(\frac{fL}{D} + K \right) \left(\frac{V^2}{2g} \right) \quad (3.2)$$

where h = head loss, ft (m); f = Darcy friction factor; L = pipe length, ft (m); D = pipe inside diameter, ft (m); V = fluid velocity, ft/s (m/s); g = gravitational constant (32.17 ft/s² or 9.807 m/s²); K = fitting loss coefficient.

The Darcy-Weisbach equation expressed in psi (Pa) is as follows:

$$\begin{aligned} P &= \frac{fL}{D \rho A^2} \frac{m^2}{(1.2 \times 10^{11})} \\ &= \frac{fL}{D \rho D^4} \frac{m^2}{(7.4 \times 10^{10})} \end{aligned} \quad (3.3)$$

where P = pressure drop, psi (Pa); m = mass flow rate, lbm/h (kg/s); ρ = fluid density, lbm/ft³ (kg/m³); A = flow area cross section, ft² (m²).

For SI units, the Darcy-Weisbach equation is as follows:

$$P = \frac{fL}{D \rho A^2} \frac{m^2}{(7.4 \times 10^{10})} \quad (3.4)$$

For both sides, pressure drop is proportional to density and to the flow rate (or velocity) squared.

This requires a known pressure drop with corresponding operating conditions to predict operation at new conditions. Note that this normalizing technique works only when there is no change in the flow regime or in shell-side flow partitioning (more or less leakage or bypassing).

Because shell-side flow is complex, do not over extrapolate. The Darcy friction factor can be determined from the graph in [Figure 3.6](#). See [Figure 3.7](#).

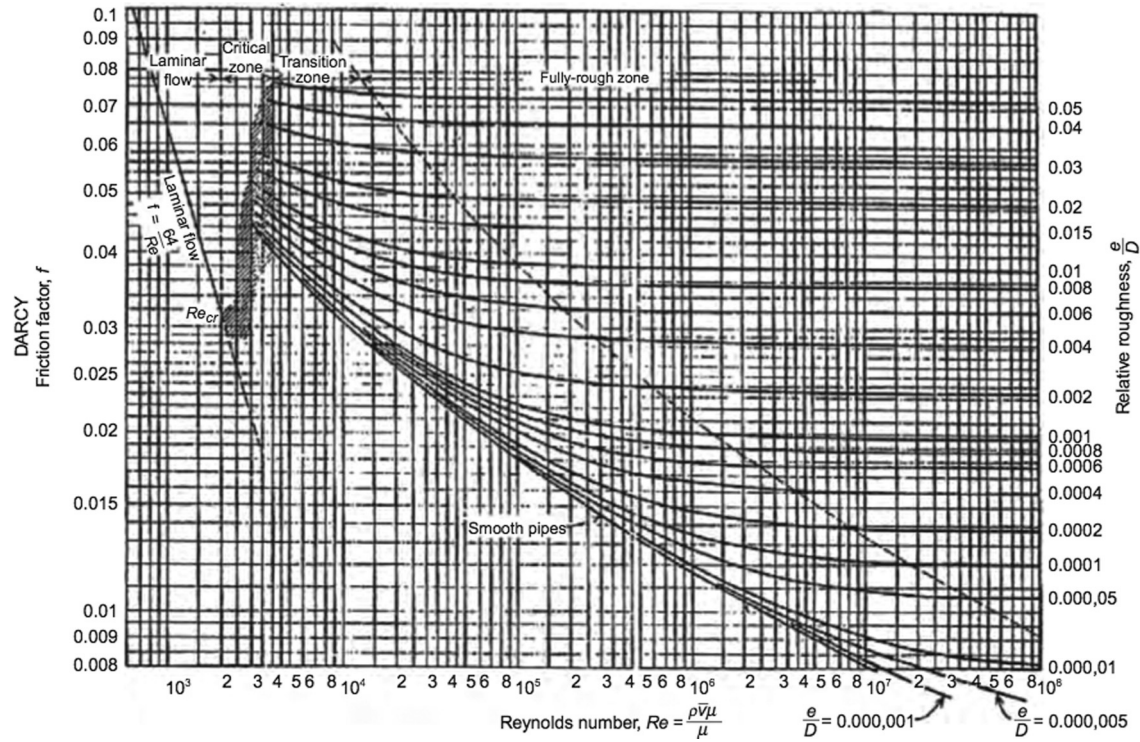


Figure 3.6 Friction factor for fully developed flow in circular pipes.

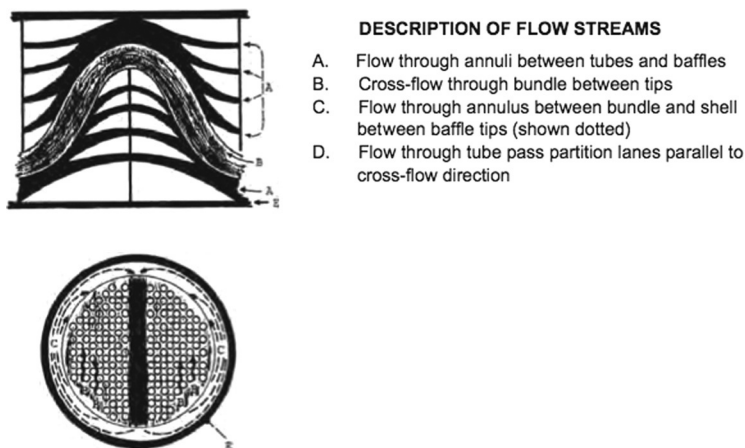


Figure 3.7 Shell-side flow stream.



3.5 BASIC HEAT TRANSFER THEORY

Sensible heat is energy transferred solely as a result of a temperature change.

3.5.1 Heat Transfer Mechanisms

Conduction

Conduction is the transfer of heat from one molecule to an adjacent molecule while the particles remain in fixed positions relative to each other. It is the primary mechanism in solids and some fluids that are stagnant or have low flow rates.

Convection

Convection is the transfer of heat by the physical movement of molecules from place to place, for example, the mixing of warmer and cooler portions of a fluid in a heater. It is the primary mechanism in fluids.

Radiation

Radiation is the process where heat waves are emitted that may be absorbed, reflected, or transmitted through a colder body. Sun heats Earth by electromagnetic waves. Hot bodies emit heat waves.

In general, most upstream facilities use conduction, convection, or a combination of the two.

Radiant energy from a direct flame, in fluid-fluid exchangers, the temperatures are not hot enough for radiation to be a significant mechanism.

It is important in calculating the heat given off by a flare. Refer to API 521 “Guidelines for Flare System Sizing and Radiation Calculation.”

3.5.2 Basic Equations

Conduction

$$q = k(A/L)(\Delta t_m) \quad (3.5)$$

Convection

$$q = hA(\Delta t_m) \quad (3.6)$$

Radiation

$$q = \sigma AT^4 \quad (3.7)$$

where q = heat transfer rate, Btu/h; A = heat transfer area, ft²; Δt_m = temperature difference, °F; k = thermal conductivity, Btu/h ft °F; h = film coefficient, Btu/h ft² °F; L = distance heat energy is conducted, ft; σ = Stefan-Boltzmann constant 0.173×10^3 Btu/h ft.

The value of “ h ” is a proportionality constant used to characterize liquid film resistance.

The value “ k ” is the thermal conductivity of the solid separating the two fluids.

3.5.3 Flow of Heat

If heat is being transferred through various layers and via various modes, two conclusions can be drawn regarding the flow of heat:

- The heat flow (q) is equal through all layers, thus

$$q_1 = q_2 = q_3 \quad (3.8)$$

- The heat flow is equal to the overall temperature difference divided by the total thermal resistance, thus

$$q = \frac{\Delta t_M}{\sum R} \quad (3.9)$$

where

$$\begin{aligned}
 R &= \text{thermal resistance of each layer, h/}^{\circ}\text{F/Btu} \\
 &= \frac{L}{kA} (\text{conduction}) \\
 &= \frac{1}{hA} (\text{convection})
 \end{aligned}$$

If $\sum R$ is defined as $\left(\frac{1}{UA}\right)$ Equation (3.9) can be rewritten as

$$q = UA\Delta t_M \quad (3.10)$$

where U = overall heat transfer coefficient, Btu/h ft² °F.

Equation (3.10) is the basic equation used for heat transfer calculations. Equation (3.10) can be rearranged to solve for the area required, which is the major design parameter for all heat exchanger design calculations.

$$A = \frac{q}{U\Delta t_m} \quad (3.11)$$

To calculate the area, three parameters must be determined:

- Mean temperature difference, Δt_M
- Heat transfer coefficient, U
- Heat duty, q

3.5.4 Multiple Transfer Mechanisms

Heat exchangers transfer energy in three steps

- Convective steps
 - Hot fluids to exchanger tube
 - Exchanger tube to cold fluid
- Conductive step
 - Flow of heat through exchanger tube wall



3.6 DETERMINATION OF MEAN TEMPERATURE DIFFERENCE

3.6.1 Mean Temperature Difference

The driving force for heat transfer is the temperature differences between the two fluid streams.

Because the temperature of the process fluid changes as it flows, a mean temperature difference (MTD) must be used.

Mean temperature difference

$$\text{MTD} = F(\text{LMTD}) \quad (3.12)$$

where

$$\text{LMTD} = [(\Delta t_1 - \Delta t_2) / \ln(\Delta t_1 / \Delta t_2)] \quad (3.13)$$

where Δt_1 = largest terminal Δt ; Δt_2 = smallest terminal Δt ; \ln = logarithm to the base e ; F = correction factor for heat exchanger geometry, =1.0 for pipe-in-pipe and counter-current, =1.0 for boiling or condensing of pure components.

For exchangers with straight-line heat release curves, the MTD is equal to the calculated log mean temperature difference (LMTD) with the “ F ” correction factor equal to 1. Single- and two-phase exchangers and some condensers typically fall into this category. LMTD is shown in [Figure 3.8](#).

The expression for the F -correction factor is used for a single TEMA E-shell with an even number of tube passes.

Two fluids may transfer heat in either:

- Counter-current direction ([Figure 3.9](#)) or
- Co-current direction ([Figure 3.10](#))

The relative direction of the two fluids influences the value of the LMTD and thus, the area required to transfer a given amount of heat.

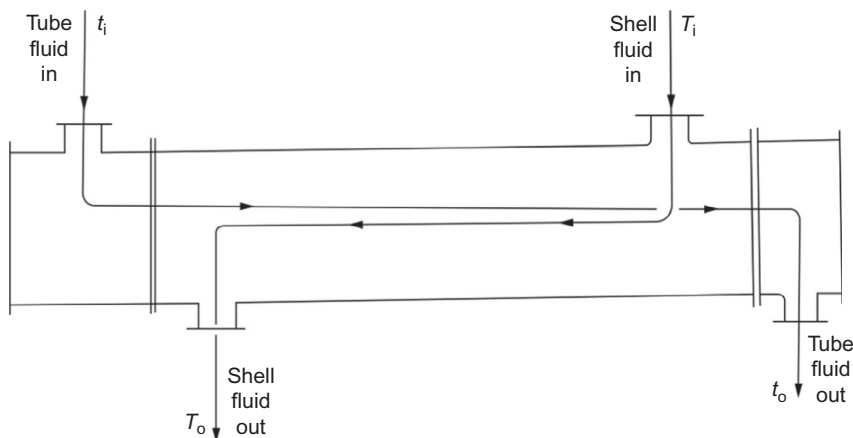


Figure 3.8 Log mean temperature difference for an exchanger with straight-line heat release curves.

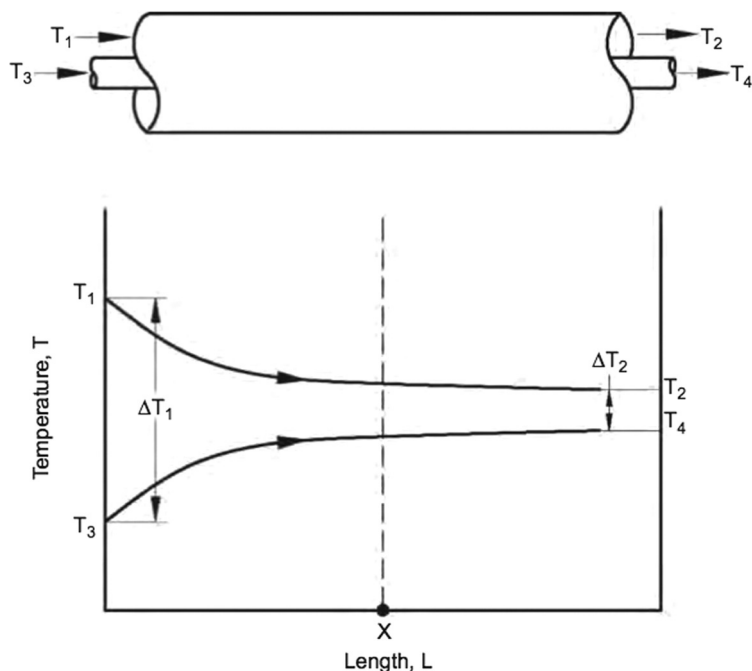


Figure 3.9 Counter-current flow (T_1 , hot fluid in; T_2 , hot fluid out; T_3 , cold fluid in; T_4 , cold fluid out).

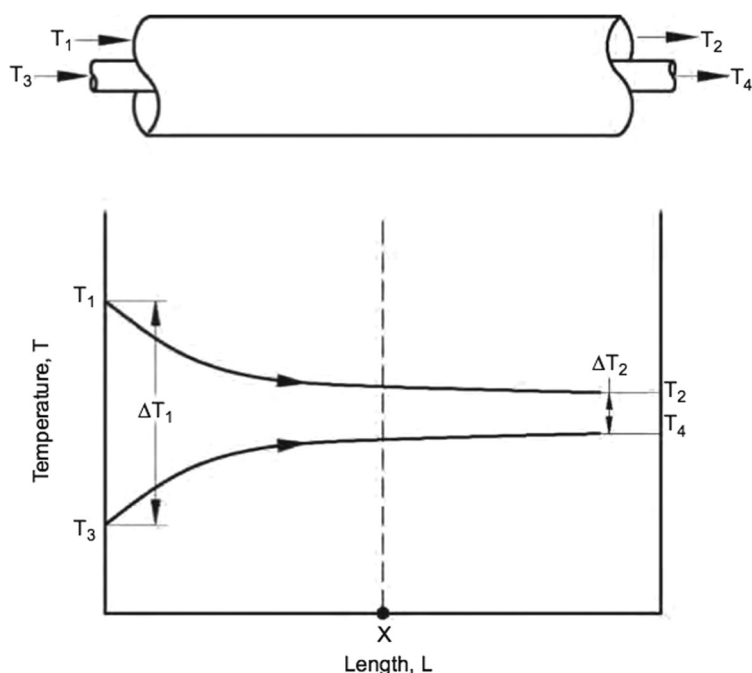


Figure 3.10 Co-current flow (T_1 , hot fluid in; T_2 , hot fluid out; T_3 , cold fluid in; T_4 , cold fluid out).

Example 3.1: Log Mean Temperature Difference

Determination

Given

A hot fluid enters a concentric pipe at a temperature of 300 °F and is to be cooled to 200 °F by a cold fluid entering at 100 °F and heated to 150 °F.

Determine

LMTD for

- (a) co-current flow
- (b) counter-current flow

Solution

(a) Co-current flow

Side	Hot Fluid	Cold Fluid	ΔT
Hot fluid inlet	300	100	200
Hot fluid outlet	200	150	50
$\Delta t_1 = \text{GTTD} = 200$; $\Delta t_2 = \text{LTDD} = 50$			

Thus,

$$\begin{aligned} \text{LMTD} &= \frac{\Delta t_1 - \Delta t_2}{\ln \left(\frac{\Delta t_1}{\Delta t_2} \right)} = \frac{200 - 50}{\ln \frac{200}{50}} \\ &= 108.2^\circ\text{F} \end{aligned}$$

(b) Counter-current flow

Side	Hot Fluid	Cold Fluid	ΔT
Hot fluid inlet	300	150	150
Hot fluid outlet	200	100	100
$\Delta t_1 = \text{GTTD} = 150$; $\Delta t_2 = \text{LTDD} = 100$			

Thus,

$$\begin{aligned} \text{LMTD} &= \frac{\Delta t_1 - \Delta t_2}{\ln \left(\frac{\Delta t_1}{\Delta t_2} \right)} = \frac{150 - 100}{\ln \frac{150}{100}} \\ &= 123.3^\circ\text{F} \end{aligned}$$

Figure 3.11 is a graphical solution for determining LMTD.

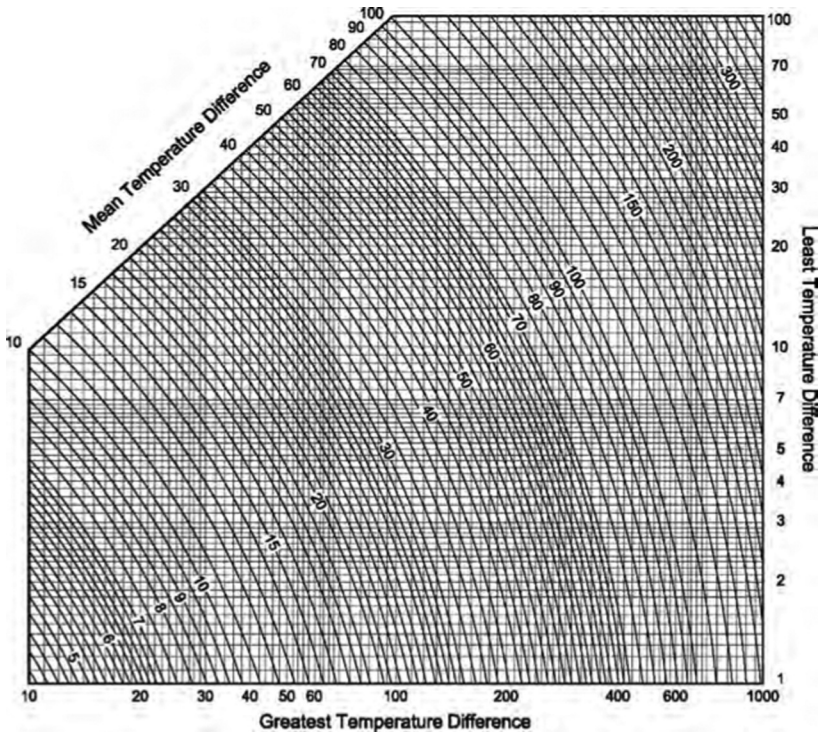


Figure 3.11 Graph for solving LMTD.

3.6.2 Log Mean Temperature Difference

The assumptions used in the derivation of LMTD are listed below:

- Constant “ U ” throughout exchanger
- Smooth-line temperature profiles
- True co-current or counter-current flow

These assumptions are not generally valid for the majority of heat exchangers, and corrections must be made.

3.6.3 Nonconstant “ U ”

The overall heat transfer coefficient will not remain constant. Calculations based on a value of U taken midway between ends of the exchanger are usually accurate enough. If there is a considerable variation in U from one end of the exchanger to the other, then a step-by-step numerical integration is necessary.

3.6.4 Different Flow Arrangements

More complicated flow arrangements than simple co-current and counter-current exist in heat exchanger equipment. These configurations are more difficult to treat analytically, thus a temperature correction factor is introduced, which is a function of:

- Exchanger inlet and outlet temperature
- Type flow arrangements for both streams

Correction factors are given in TEMA and in the Gas Processors Suppliers Association (GPSA) Engineering Data Book. As a practical matter, an exchanger with a correction factor < 0.8 should not be used. Additional data on correction factors are presented in the shell-and-tube and air cooler sub-sections of this chapter.

3.6.5 Nonlinear Temperature Profile

Nonlinear temperature profiles occur in three situations (Figure 3.12)

- Condensation plus subcooling
- Vaporization plus subcooling
- Curved T plot versus Q plot

For purposes of analysis, this process may be considered the super position of two or more exchangers as was depicted in Figure 3.10. The total heat

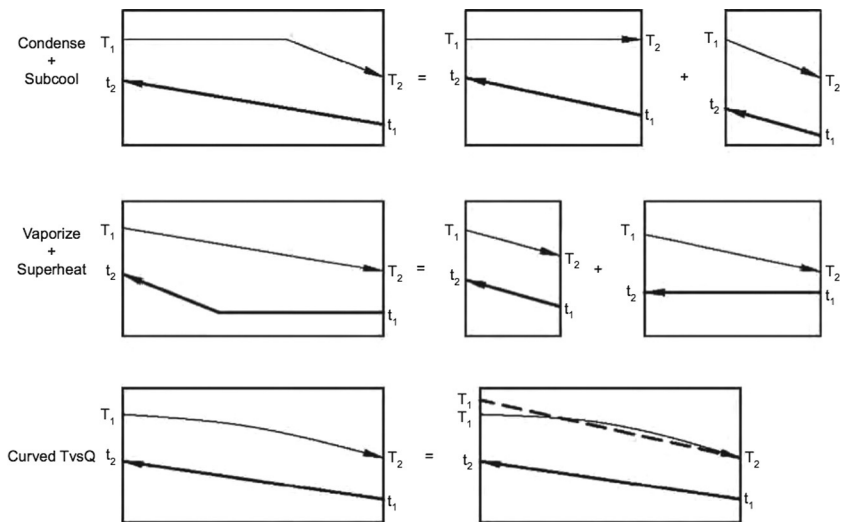


Figure 3.12 Nonlinear temperature profiles.

exchanger area cannot be calculated using $Aq/(u) \text{ DtM}$ when temperature profiles are nonlinear. Summaries of several procedures follow:

1. Calculate the area for each segment of the exchanger. The heat transfer coefficient (U) varies through the exchanger. The correct U for each segment depends on the fluid phase in that segment. The exchanger area is the sum of the segments' areas.
2. If an overall heat transfer coefficient is known, the area can be calculated using a weighted LMTD (WLMTD), expressed as follows:

$$\text{WLMTD} = \frac{\sum q}{\frac{q_1}{\text{LMTD}_1} + \frac{q_2}{\text{LMTD}_2} + \frac{q_3}{\text{LMTD}_3}} \quad (3.14)$$

Example 3.2: Application of WLMTD

Given: Figure 3.13

Determine

Cooling and condensation weighted temperature difference.

Solution

Sensible $q_d = 40$

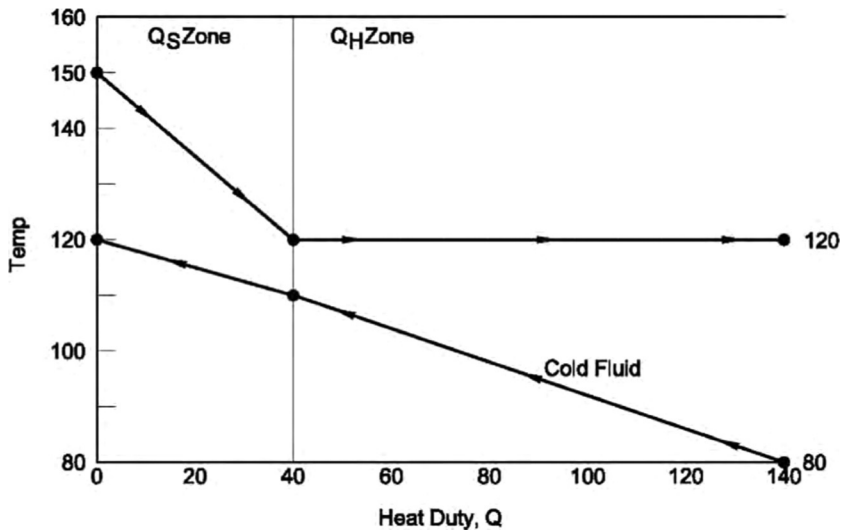


Figure 3.13 LMTD for cooling and condensation.

$$\begin{aligned}\text{LMTD}_s &= \frac{30 - 10}{\ln(30/10)} \\ &= 18.2^\circ\text{F}\end{aligned}$$

$$\text{Latent } q_h = 140$$

$$\begin{aligned}\text{LMTD}_h &= \frac{40 - 10}{\ln(40/10)} \\ &= 21.6^\circ\text{F}\end{aligned}$$

Thus,

$$\begin{aligned}\text{WLMTD} &= \frac{140}{\frac{40}{18.2} + \frac{100}{21.6}} \\ &= 20.5^\circ\text{F}\end{aligned}$$

3. If the T plot versus Q plot is curved, a linear approximation can be constructed for an LMTD calculation instead of using the divided segment approach.



3.7 SELECTION OF TEMPERATURE APPROACH (ΔT_2)

The approach temperature is an economic choice as its specification governs exchanger cost. As “ Δt_2 ” gets smaller the

- LMTD becomes smaller
- Area required becomes larger

Because the cost of an exchanger is a direct function of area, specification of approach temperature has a direct effect on cost. When specifying heat exchangers, it is usually beneficial to specify a maximum or minimum approach to the vendor. This establishes an upper or lower limit, below or above which the actual approach must occur. Common approaches are:

- Aerial coolers (18–45 °F)
- Water cooling of hydrocarbon liquids and gases (14–22 °F)
- Liquid-liquid heat exchange (20–45 °F)
- Refrigeration chillers on gas-liquid streams (7–11 °F)



3.8 DETERMINATION OF HEAT TRANSFER COEFFICIENT

3.8.1 Overview

Heat transfer coefficient for a heat exchanger can be defined as the sum of thermal resistances per unit area as shown by the following equation:

$$U = \frac{1}{\sum R_i A}$$

where U = overall heat transfer coefficient Btu/h-ft² °F.

3.8.2 Area Basis

Nearly all heat exchanger problems involve heat transfer through three or more layers around circular tubes. Typically, there are two areas to consider:

- Inside tube area (A_i)
- Outside tube area (A_o)

Unless otherwise stated, all heat transfer values presented in the literature are based on outside tube area (A_o).

3.8.3 Heat Transfer Coefficient—Clean Tube

Heat transfers through a clean tube are shown in [Figures 3.14](#) and [3.15](#). Equation (3.10), the general heat transfer equation, can be rewritten using the thermal resistances in the three layers

$$\begin{aligned} R_1 &= \text{Inside convection resistance} = \frac{1}{h_i A_i} \\ R_2 &= \text{Conduction through tube wall} = \frac{L}{k A_w} \\ R_3 &= \text{Outside convection resistance} = \frac{1}{h_o A_o} \end{aligned}$$

Thus,

$$\begin{aligned} q &= \frac{\Delta t_m}{\sum R} \\ &= \frac{\Delta t_m}{R_1 + R_2 + R_3} \\ &= \frac{\Delta t_m}{\frac{1}{h_i A_i} + \frac{L}{k A_w} + \frac{1}{h_o A_o}} \end{aligned} \tag{3.15}$$

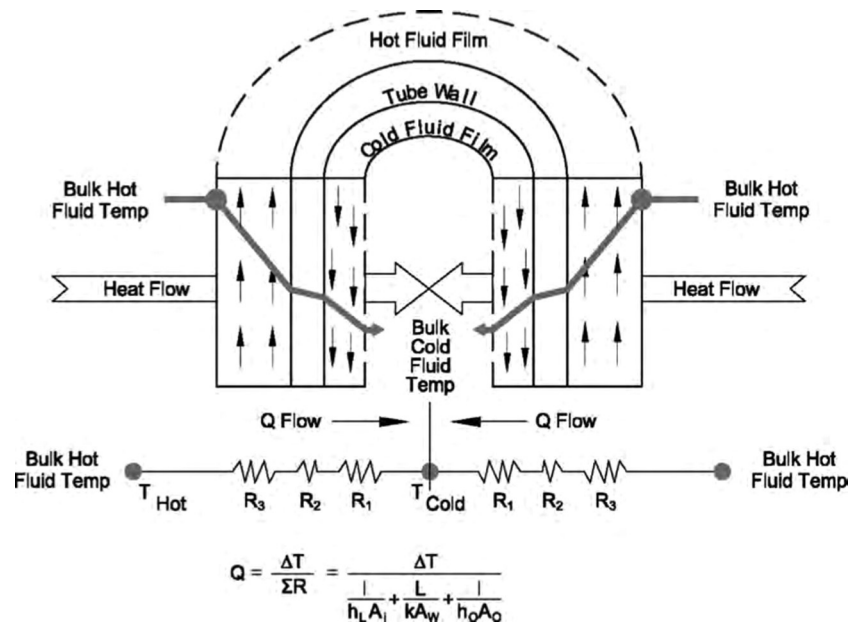


Figure 3.14 Heat transfer through a clean tube.

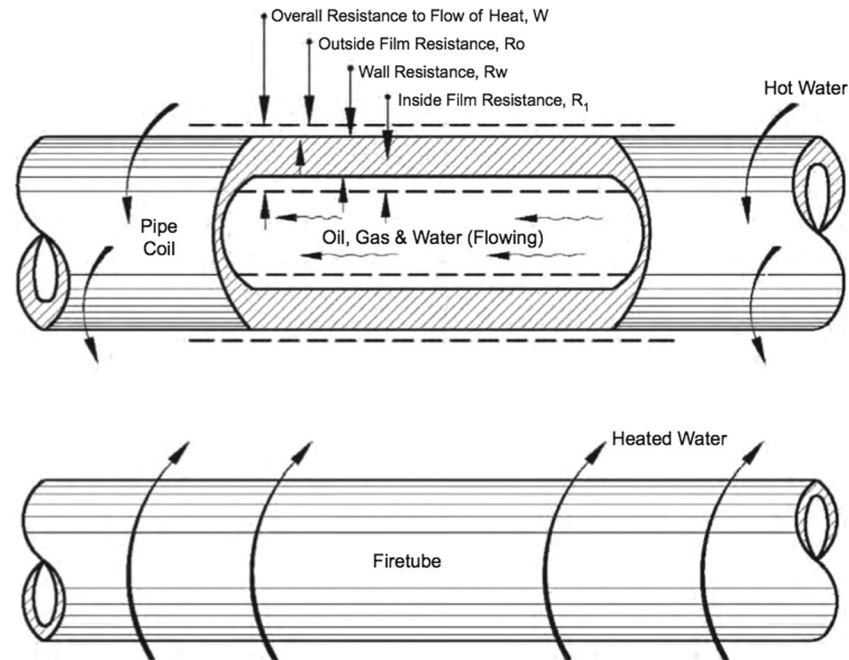


Figure 3.15 Heat transfer path involving resistances in series.

Assuming A_w (average tube area) equals A_o (outside tube wall area), Equation (3.15) can be arranged as follows:

$$q = \left[\frac{1}{\frac{1}{h_i} \left(\frac{A_o}{A_i} \right) + \frac{L}{k} + \frac{1}{h_o}} \right] A_o \Delta t_m$$

$$U = \frac{1}{\frac{1}{h_i} \left(\frac{A_o}{A_i} \right) + \frac{L}{k} + \frac{1}{h_o}}$$

where h_i = inside film coefficient, Btu/h ft² °F; h_o = outside film coefficient, Btu/h ft² °F.

3.8.4 Heat Transfer Coefficient—Fouled Tube

Thin layers of dirt, scale, corrosion products, or degradation products can build up over time on the inside or outside of the tube and reduce the rate of heat transfer (Figure 3.16).

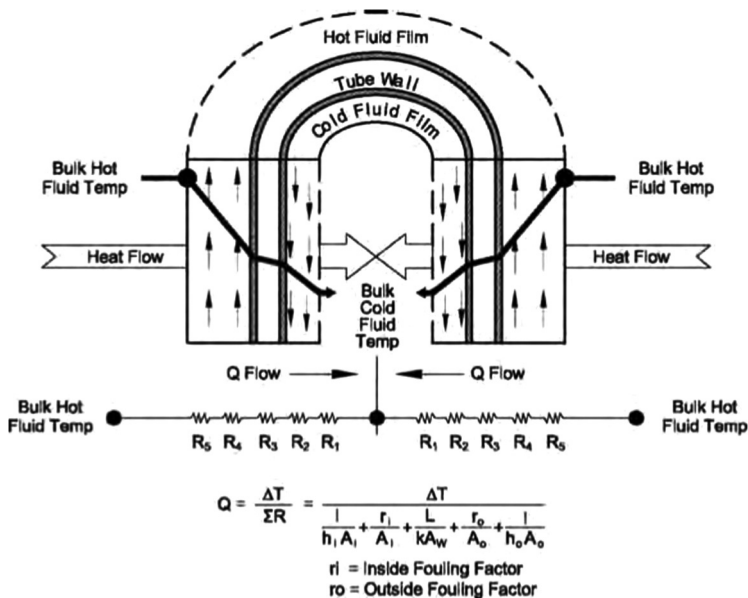


Figure 3.16 Heat transfer through fouled tubes.

3.8.4.1 Fouling Factors

Fouling factors are included in calculations to account for film buildup. The resistance built up on the outside is designated r_o , and the resistance built up on the inside is designated r_i . They are presented in the literature for various materials and conditions.

Because scale or dirt resistance increases with time in service, some time basis must be chosen for numerical values of fouling factors. The time basis of 1 year is commonly used.

Five thermal resistances must be considered when calculating the heat transfer coefficient for fouled tubes:

$$\begin{aligned}
 R_1 &= \text{Inside convection resistance} = \frac{1}{h_i A_i} \\
 R_2 &= \text{Inside fouling resistance} = \frac{r_i}{A_i} \\
 R_3 &= \text{Conduction through wall} = \frac{L}{k A_w} \\
 R_4 &= \text{Outside fouling resistance} = \frac{r_o}{A_o} \\
 R_5 &= \text{Outside convection resistance} = \frac{1}{h_o A_o}
 \end{aligned}$$

Thus,

$$\begin{aligned}
 q &= \frac{\Delta t_m}{\sum R} \\
 &= \frac{\Delta t_m}{\frac{1}{h_i A_i} + \frac{r_i}{A_i} + \frac{L}{k A_w} + \frac{r_o}{A_o} + \frac{1}{h_o A_o}}
 \end{aligned} \tag{3.16}$$

Assuming $A_w = A_o$, then,

$$q = \left[\frac{1}{\frac{1}{h_i} \left(\frac{A_o}{A_i} \right) + r_i \left(\frac{A_o}{A_i} \right) + \frac{L}{k} + r_o + \frac{1}{h_o}} \right] A_o \Delta t_m \tag{3.17}$$

where

$$U = \frac{1}{\frac{1}{h_i} \left(\frac{A_o}{A_i} \right) + r_i \left(\frac{A_o}{A_i} \right) + \frac{L}{k} + r_o + \frac{1}{h_o}} \tag{3.18}$$

where h_i , h_o =inside, outside film coefficient, Btu/h-ft²°F; k =thermal conductivity, Btu/h ft²°F; A_i , A_o =inside, outside surface area, ft²/ft; r_i , r_o =inside, outside fouling resistance, h-ft²°F/Btu; L =wall thickness, ft.

A heat transfer coefficient based on fouled tubes can be determined by knowing the clean heat transfer coefficient and the fouling factor as show by the following equation:

$$U_{\text{fouled}} = \frac{1}{\frac{1}{U_{\text{clean}}} + r_i \left(\frac{A_o}{A_i} \right) + r_o}$$

Actual fouling in heat exchangers depends on several characteristics:

- Nature of fluid and material deposited
- Fluid velocity
- Temperature of fluid
- Temperature of tube wall
- Tube wall material
- Tube wall surfaces (finned or smooth)
- Time since last cleaning

Table 3.1 lists typical fouling factors. More detailed information can be found in TEMA Standards. Fouling information is highly proprietary. r_o plus r_i , are commonly assumed to be =0.003 h-ft²°F/Btu.

3.8.4.2 Fouling Considerations

One must not confuse real fouling with a calculation result. Fouling is material accumulation on heat transfer surfaces. The fouling factor is a:

Table 3.1 Typical Fouling Factors

Fluid	$R_o + R_i$ (h ft ² °F/Btu)
Crude oil	0.002
Natural gas	0.001
Overhead products	0.001
Lean oil	0.002
Rich oil	0.002
LPG	0.001
Amine solutions	0.002
Acid gas	0.001
Refrigerant liquids	0.001
Engine exhaust	0.010
Cooling water	0.001

- Calculated result derived from plant data
- Fudge factor
- Number from a table

Typically, one cannot determine how much fouling is shell-side fouling.

3.8.4.3 Fouling Mechanisms

Use trend analysis to help pinpoint the type of fouling. Sources of fouling include:

- Particulate or solids deposition
- Salt crystallization or insolubility
- Polymerization, coking, or reaction to solid
- Human additives
- Biological growth
- Corrosion

Methods to prevent or minimize fouling include:

- Particulate or solids deposition—maintain sufficient velocity to keep from settling
- Salt crystallization or insolubility—manage concentration, skin and bulk temperatures, and pH
- Polymerization, coking, or reaction to solid—keep temperature below activation or control exponential reaction rate
 - Chemical additives—ensure that they are always below their boiling or decomposition temperature
 - Biological growth—remove or control rate of deposition with appropriate chemical additives

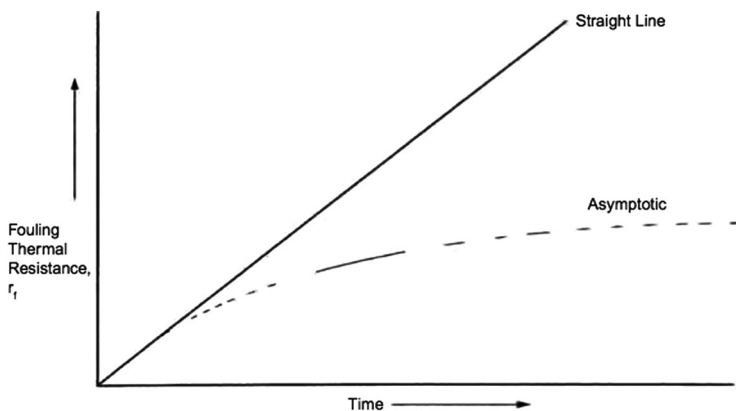


Figure 3.17 Fouling mechanisms.

- Corrosion—adjust the process to prevent or select appropriate material

Figure 3.17 is a graph displaying three different fouling trends that can be used to help identify the cause of fouling.

3.8.5 Evaluating Performance

The performance of a heat exchanger can be evaluated by the following equations:

$$R_f = \frac{1}{U_{\text{observed}}} - \frac{1}{U_{\text{expected}}}$$

where

$$\frac{1}{U_{\text{expected}}} = \frac{1}{U_{\text{clean}}} + R_{f(\text{expected})}$$

Questions that need to be addressed include:

- What is time trend?
- What is HEX fouling?

For design

1. First $U_{\text{expected}}/U = 125\%$
2. Then $R_f = 1/U - 1/U_{\text{expected}}$
3. Exceptions
 - Will this HEX foul?

Figure 3.18 illustrates fouling nomenclature.

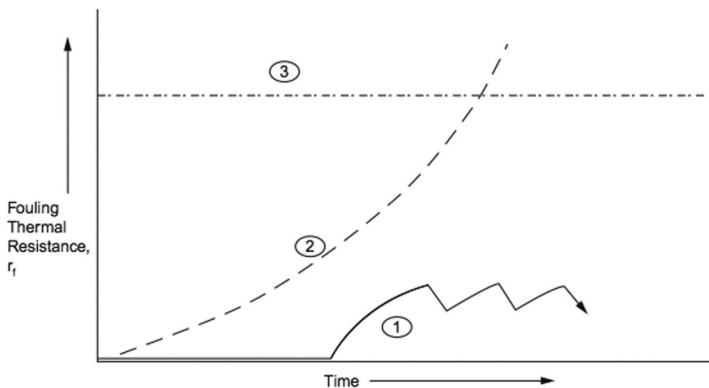


Figure 3.18 Fouling data trends nomenclature.

3.8.6 Heat Transfer Research Inc. Computer Simulation Programs

A computer simulation program is the workhorse for the heat exchanger suite, and

- Is relatively easy to set up
- Used for design and troubleshooting
- Handles many service types, including
 - Air coolers
 - Sensible heat transfer
 - Single phase, except for “two phase” with <70% duty as latent heat
 - Condensing
 - Tube-side and shell-side boiling
 - Vertical thermo-siphons
 - Horizontal thermo-siphons
 - Kettle re-boilers
 - Plate-and-frame heat exchangers

Fouling is typically rapid and dramatically limits heat transfer. See [Table 3.2](#).

Table 3.2 Services Susceptible to Fouling		
Problem	Susceptible Service	Avoid By
Participate or solids deposition	Any stream—liquid, gas, or two phase	Maintain sufficient velocity to keep from settling
Salt crystallization or insolubility	Stream containing water phase	Manage concentration, skin and bulk temperatures, and pH
Polymerization, coking, or reaction to solid	Heavy hydrocarbon liquids and streams contaminated with catalyst	Keep temperature below activation or control exponential reaction rate
Man-made additives	Any liquid or two-phase streams	Avoid
Biological growth	Water or water-contaminated streams	Kill or control rate of growth. This is materials, temperature, and velocity dependent
Corrosion	Practically any stream	Adjust process to prevent corrosion and select appropriate materials to resist corrosion



3.9 CALCULATION OF FILM COEFFICIENTS

Calculation of heat transfer film coefficients often require trial-and-error calculations and are time consuming to do by hand. Computers are often used to make detailed calculations. If calculations are done by hand, care should be taken to ensure that all units are consistent.

Most liquids and gases are poor conductors of heat. The film of fluid next to the tube wall generally causes the controlling resistance to heat transfer. Increasing the velocity of the fluid stream can reduce the thickness of the film.

Typical U values can be used for rough or preliminary calculations. They are presented throughout the remainder of this chapter.

Film coefficients depend on service. Available tools include equations, charts, simplifying assumptions, and computer programs. For all but the simplest cases, use a computer program. For quick estimates, use [Table 3.3](#) for approximate film coefficients. For sensible streams, calculating film coefficients is straightforward.

Table 3.3 Approximate Film Coefficients

Service or Fluid	Shell or Tube Side Film Coefficient Btu/h °F ft ² (Based on Bare Outside Area)
Sensible	
Pure water	1400
HC, 0.5 cP	400
HC, 2.0 cP	250
HC, 10 cP	150
Gases	
Light HC, 150 psig	100
Air, 10 psig	15
Air, 300 psig	60
Condensing	
Steam	1000
Light HC	200
Heavy HC	100
Subcooling	50
Boiling	
Water	1000
Light HC	300
Heavy HC	150
Air cooled (fin fan)	
Air side	175

If economic sizing criteria are used, velocity or pressure gradient is optimized and equations can be solved.

- Economic sizing criteria
 - Equalize or balance film (adjusted) coefficients
- For new designs, balance pumping or compression cost against cost of heat transfer surface area
- Velocity considerations
 - 0.4–0.6 psi/axial ft (9048–13,570 Pa/axial m) shell side \approx 2–3 ft/s for liquids (0.61–0.91 m/s)
 - 0.2–0.3 psi/ft (4524–6784 Pa/m) tube side [3/4-in. tubes (19.05 mm)] \cong 7–9 ft/s for liquids (2.1–2.7 m/s)

Expensive metallurgy suggests design at a higher velocity. Active fouling (contaminated streams) suggests design at a higher velocity. The designer must consider a turndown operation and keep the pressure gradient above the particulate fouling threshold;

- psi/axial ft (2262 Pa/axial m) shell side \cong 1 ft/s (0.305 m/s)
- 0.05 psi/ft (1131 Pa/m) tube side \cong 4 ft/s (1.22 m/s)

The designer should use the economic sizing criteria for single-phase, two-phase, multicomponent condensing, and flow boiling.

Pure component condensing and boiling typically have very large film coefficients, and accuracy is not important. Multicomponent condensing and boiling, as well as a combination of sensible plus latent heat transfer, are very complicated.

These should be analyzed using an appropriate computer program. For boiling streams, design and operate in the nucleate regime. For condensing streams, the regime dramatically changes as heat is transferred. See [Figure 3.19](#).

3.9.1 Inside Film Coefficient

The inside film coefficient represents the resistance to heat flow caused by the change in flow regime from turbulent flow in the center of the tube to laminar flow at the tube surface. The inside film coefficient is calculated from the following equation:

$$h_i = \left[0.022 \left(\frac{dG}{\mu} \right)^{0.8} \left(\frac{C_p}{k} \right)^{0.4} \left(\frac{\mu}{\mu_w} \right)^{0.16} \right] \frac{k}{p} \quad (3.19)$$

where h_i = fluid film heat transfer coefficient, Btu/h ft² °F; D = tube inside diameter, ft; k = fluid thermal conductivity, Btu/h ft °F (refer to

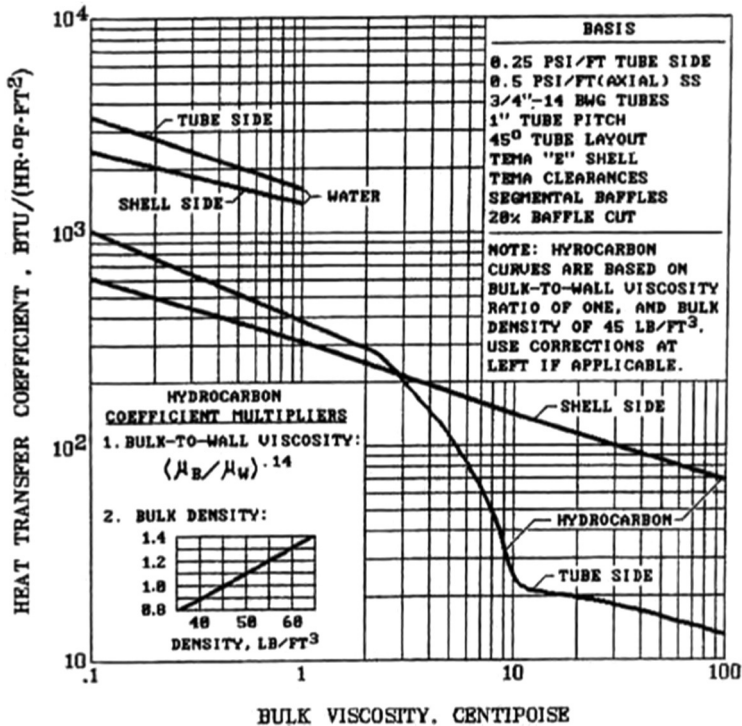


Figure 3.19 Heat transfer coefficients.

Figures 3.20–3.22; G = mass velocity of fluid, lb/h ft^2 [refer to Equations (3.16) and (3.17)], μ = fluid viscosity, lb/h ft (2.41) μ_{CP} , =refer to Figures 3.23–3.24; μ = fluid viscosity, lb/h ft ; μ_w = fluid viscosity at tube wall, lb/h ft ; C_p = fluid specific heat, $\text{Btu/lb } ^\circ\text{F}$.

3.9.2 Mass Velocity of a Fluid

Liquids

$$G = 18.6 \frac{Q_L (SG)}{D^2} \quad (3.20)$$

Gases

$$G = 4053 \frac{Q_g S}{D^2} \quad (3.21)$$

where Q_L = liquid flow rate per tube, BPD; Q_g = gas flow rate per tube, MMscfd; SG = liquid-specific gravity (relative to water); S = gas-specific gravity (relative to air); D = tube inside diameter, ft.

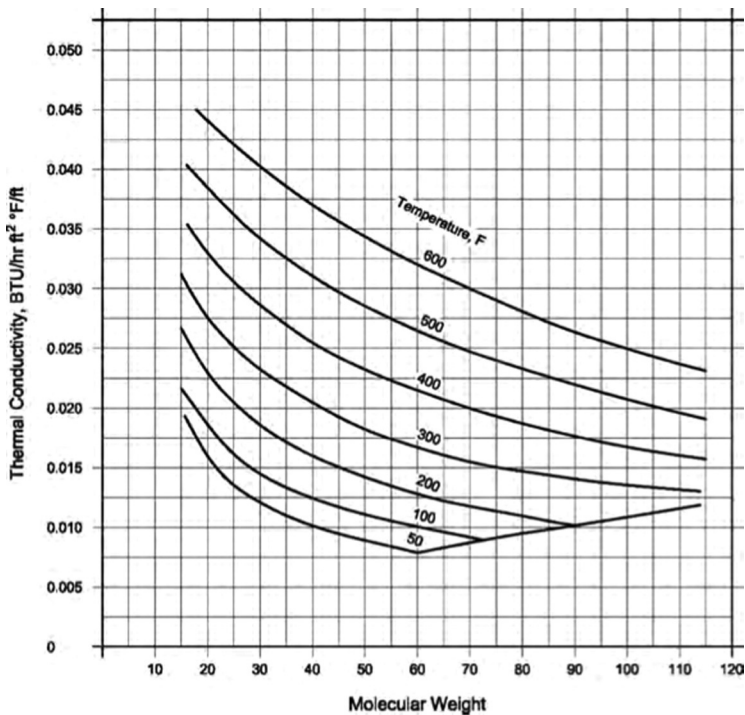
Thermal conductivity, viscosity, and specific heat of various heat medium fluids are given in [Figures 3.25–3.29](#).

3.9.3 Outside Film Coefficient in a Liquid Bath

The outside film coefficient is a result of natural or free convection. Temperature variations in the fluid cause density variations. Density variations cause the fluid to circulate, which produces free convective heat transfer. For horizontal pipes and tubes spaced more than one diameter apart, the following applies:

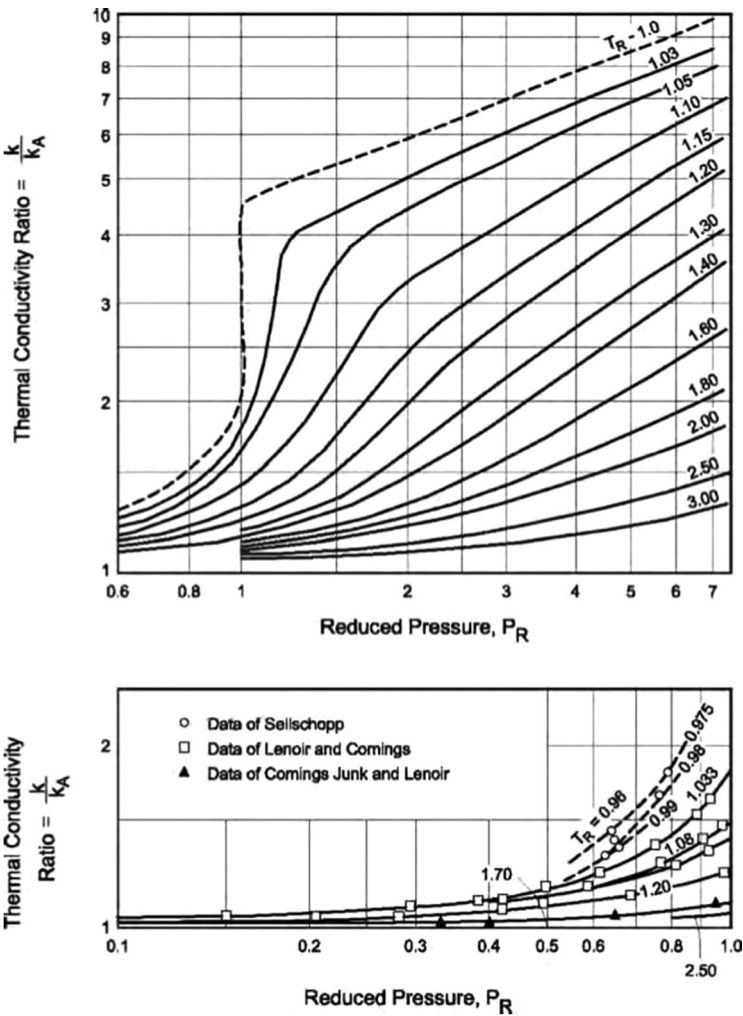
$$h_o = 116 \left[\frac{k^3 C^2 \rho \beta \Delta t_m}{\mu d_o} \right]^{0.25} \quad (3.22)$$

where h_o = outside film coefficient, Btu/h ft² °F; k = bath fluid thermal conductivity, Btu/h ft °F (refer to [Table 3.4](#)); C = bath fluid heat capacity, Btu/lb °F; ρ = bath fluid density, lb/ft³; μ = bath fluid viscosity, C_p ; Δt_m = mean



(a)

Continued



(b)

Figure 3.20—cont'd (a) Thermal conductivity of hydrocarbon gases at standard pressure (14.67 psia) (value is multiplied by the ratio k/k_A from (b)). (b) Thermal conductivity ratio for gases (value is multiplied by the thermal conductivity value from (a)).

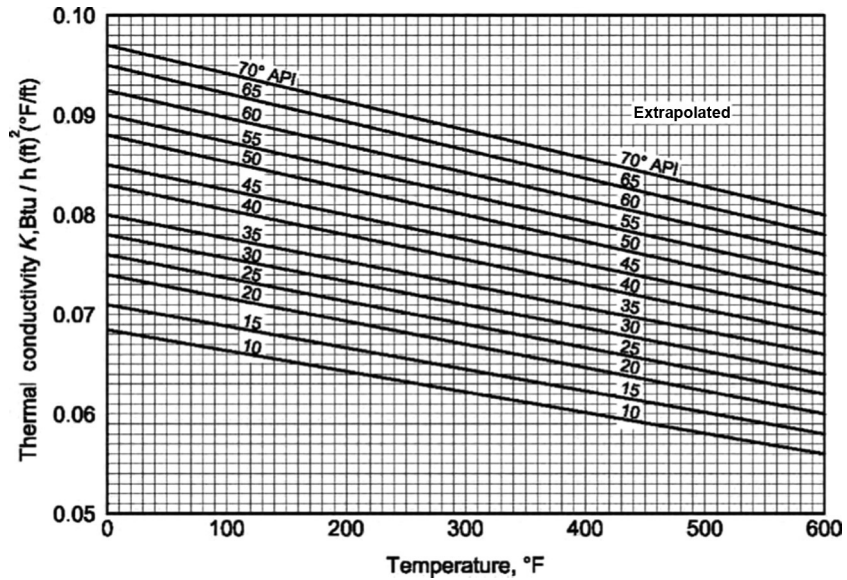


Figure 3.21 Thermal conductivity of hydrocarbon liquids.

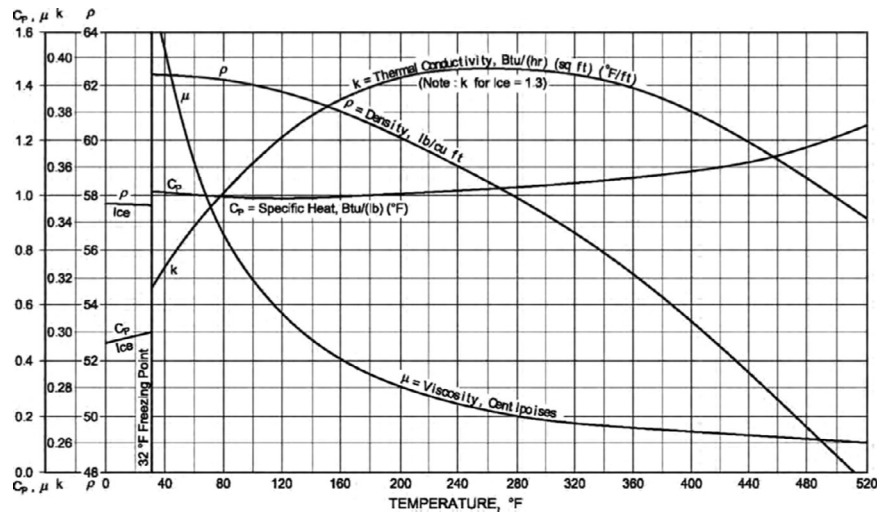


Figure 3.22 Physical properties of water.

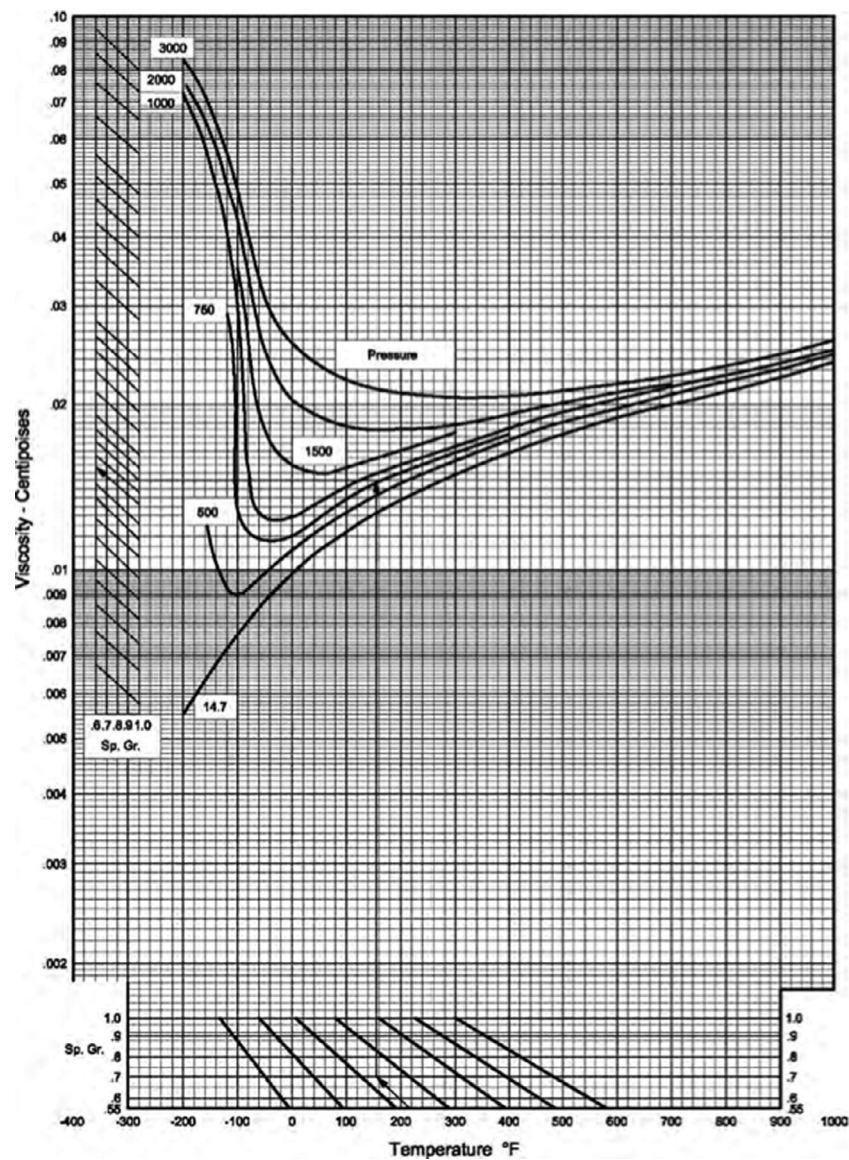


Figure 3.23 Hydrocarbon gas viscosity.

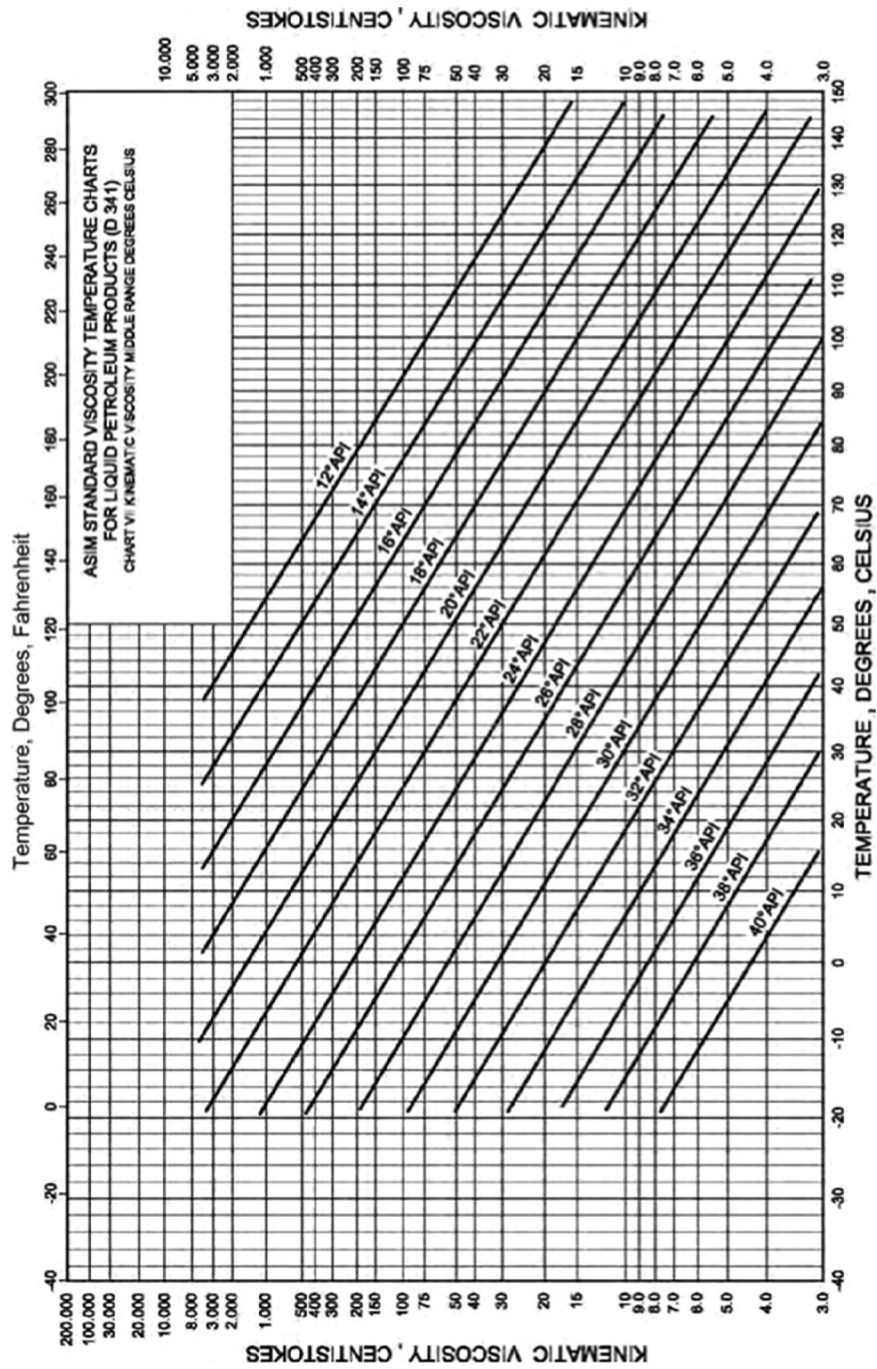


Figure 3.24 Hydrocarbon liquid viscosity.

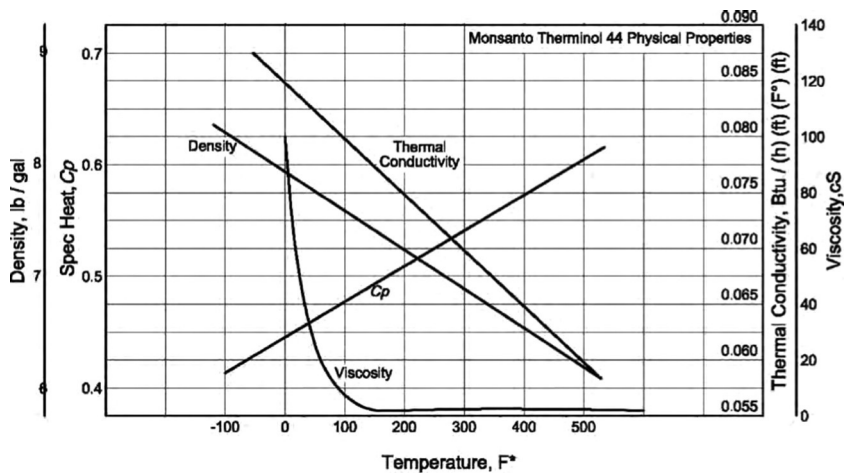


Figure 3.25 Heat transfer coefficient, Monsanto Therminol 44 (use 50-450 °F; maximum film temperature).

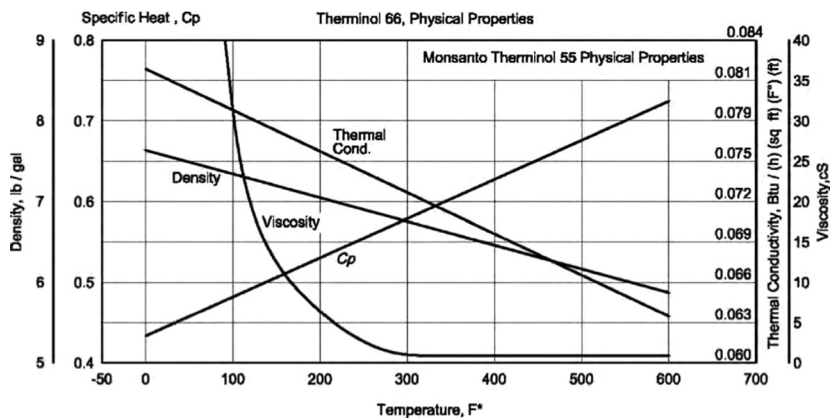


Figure 3.26 Heat transfer coefficient, Monsanto Therminol 55 (use 0-600 °F; maximum film temperature 635 °F).

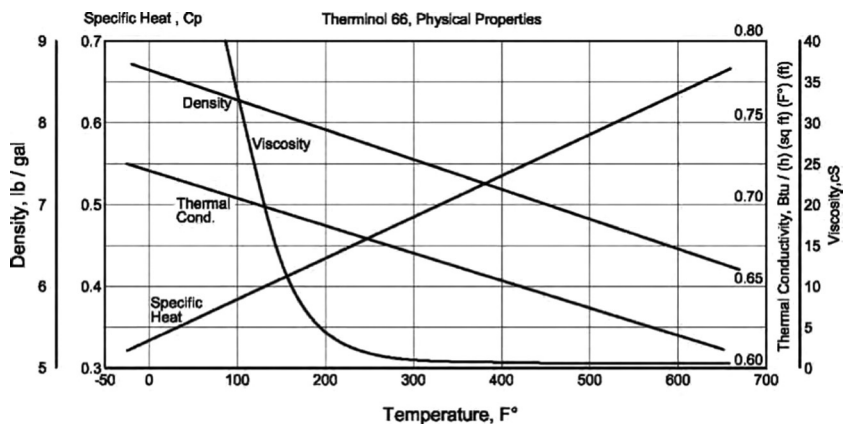


Figure 3.27 Heat transfer coefficient, Monsanto Therminol 55 (use 0-600 °F; maximum film temperature 635 °F).

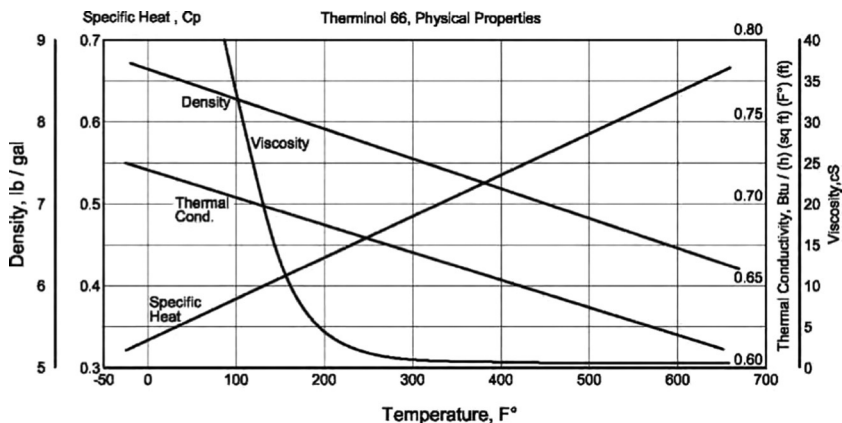


Figure 3.28 Heat transfer coefficient, Monsanto Therminal 66 (use 0-650 °F; maximum film temperature 750 °F).

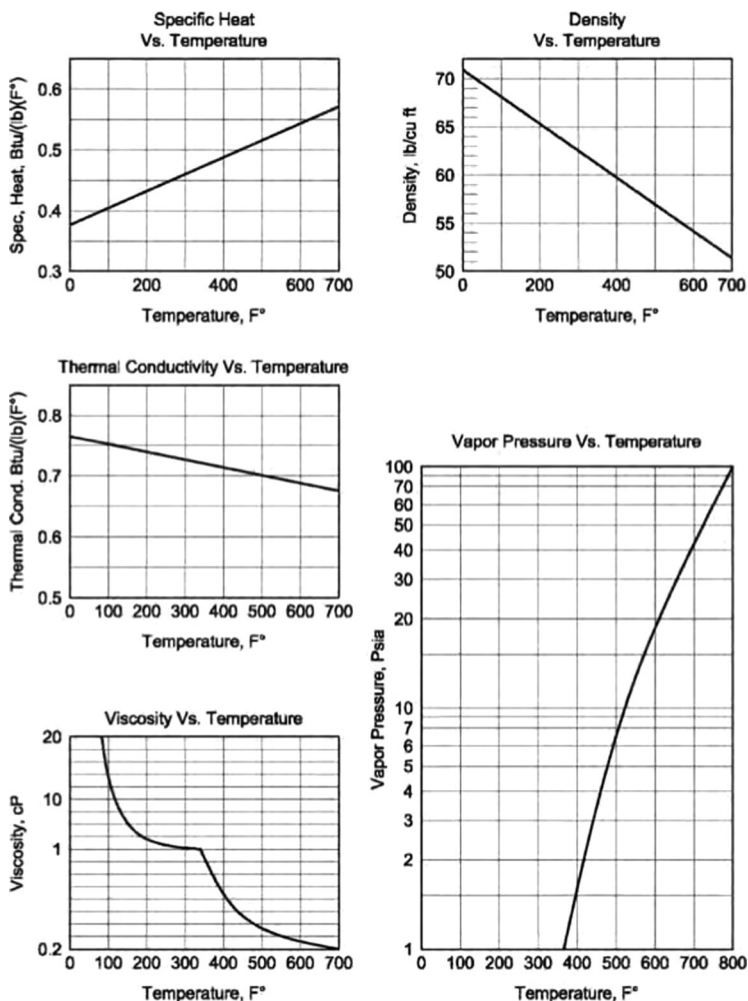


Figure 3.29 Physical properties of Dowtherm "G" versus temperature.

Table 3.4 Thermal Conductivity of Water

Temperature (°F)	Thermal Conductivity (Btu/h ft °F)
32	0.343
100	0.363
200	0.393
300	0.395
420	0.376
620	0.275

Table 3.5 Coefficients of Thermal Expansion (β)

Bath Fluid	Coefficient (1/°F)
Water	0.0024
Dowtherms	0.00043
Therminols	0.00039
Mobiltherms	0.0035

temperature difference, °F; d_o = pipe outside diameter, in.; β = bath fluid coefficient of thermal expansion, 1/°F (refer to [Table 3.5](#)).

Bath fluid coefficients of thermal expansion are given in [Table 3.5](#). The density of water is 1 divided by the specific volume given in the steam tables (see [Table 3.6](#)).

3.9.4 Outside Film Coefficient for Shell-and-Tube Exchangers

For shell-and-tube heat exchangers with shell-side baffles and shell-side fluid flow perpendicular to tubes, the outside film coefficient for a shell-and-tube exchanger is determined from:

$$h_o = 0.6K \left(\frac{C\mu_e}{k} \right)^{0.33} \left(\frac{DG_{\max}}{\mu_e} \right)^{0.6} \left(\frac{k}{D} \right) \quad (3.23)$$

where h_o = outside film coefficient, Btu/h ft² °F; D = tube outside diameter, ft; k = fluid thermal conductivity, Btu/h ft °F; G_{\max} = maximum mass velocity of fluid, lb/h ft²; C = fluid specific heat, Btu/lb °F; μ_e = fluid viscosity, lb/h ft; K = coefficient from [Table 3.7](#).

Table 3.6 Properties of Dry Saturated Steam

Time θ	Press psig P	Specific Volume			Enthalpy			Entropy			Time θ
Temp T $^{\circ}F$	P	Saturated Liquid V_{liq}	R_{vap}	Saturated Vapor V_{vap}	Saturated Vapor H_{liq}	R_{vap}	Saturated Vapor H_{vap}	Saturated Liquid He_{liq}	R_{vap}	Saturated Vapor He_{vap}	Temp $^{\circ}F$ T
32	0.08854	0.01602	3308	3208	0.00	1075.3	1075.8	0.0000	2.1877	2.1877	32
35	0.09995	0.01602	2947	2947	3.02	1074.1	1077.1	0.0061	2.1709	2.1770	35
40	0.12170	0.01602	2444	2444	8.05	1071.3	1079.3	0.0162	2.1435	2.1597	40
45	0.14752	0.01602	2036.4	2036.4	13.06	1068.4	1081.5	0.0262	2.1167	2.1429	45
50	0.17811	0.01603	1703.2	1703.2	18.07	1065.8	1083.7	0.0361	2.0903	2.1264	50
60	0.2583	0.01604	1206.8	1206.7	28.06	1059.9	1088.0	0.0555	2.0393	2.0948	60
70	0.3631	0.01605	867.8	857.9	38.04	1054.3	1092.3	0.0745	1.9902	2.0647	70
80	0.5069	0.01608	632.1	633.1	48.02	1048.6	1096.6	0.0932	1.9428	2.0380	80
90	0.5982	0.01610	468.0	468.0	57.99	1042.9	1100.9	0.1115	1.8972	2.0087	90
100	0.9492	0.01613	350.3	350.4	67.97	1037.2	1105.2	0.1295	1.3531	1.9828	100
110	1.2748	0.01617	265.3	265.4	77.94	1031.8	1109.5	0.1471	1.8106	1.9577	110
120	1.6924	0.01620	203.25	203.27	87.92	1025.8	1113.7	0.1645	1.7694	1.9339	120
130	2.2225	0.01625	157.32	157.34	97.90	1020.0	1117.9	0.1818	1.7256	1.9112	130
140	2.2886	0.01629	122.99	123.01	107.39	1014.1	1122.0	0.1984	1.6910	1.3894	140
150	3.718	0.01634	97.04	97.07	117.99	1008.2	1126.1	0.2149	1.6537	1.3685	150
160	4.741	0.01629	77.27	77.29	127.89	1002.3	1130.2	0.2311	1.6174	1.3485	160
170	5.992	0.01615	62.04	62.06	137.93	996.3	1134.2	0.2472	1.5822	1.8293	170
180	7.510	0.01651	50.21	50.23	147.92	990.2	1138.1	0.2630	1.5480	1.8109	180
190	9.339	0.01657	40.94	40.96	157.95	984.1	1142.0	0.2785	1.5147	1.7932	190
200	11.528	0.01663	33.62	23.64	167.99	977.9	1145.9	0.2938	1.4824	1.7762	200

Continued

Table 3.6 Properties of Dry Saturated Steam—cont'd

Time θ	Press psig P	Specific Volume			Enthalpy			Entropy			Time θ
		Saturated Liquid V_{liq}	R_{vap}	Saturated Vapor V_{vap}	Saturated Vapor H_{liq}	R_{vap}	Saturated Vapor H_{vap}	Saturated Liquid He_{liq}	R_{vap}	Saturated Vapor He_{vap}	
Temp T	P										Temp $^{\circ}F$ T
210	14.123	0.01670	27.80	27.32	178.05	971.6	1149.7	0.3090	1.4508	1.7398	210
212	14.696	0.01672	26.73	26.80	180.07	970.3	1150.4	0.3120	1.4446	1.7568	212
220	17.186	0.01677	23.13	22.15	188.13	965.2	1153.4	0.3239	1.4201	1.7440	220
230	20.780	0.01634	19.265	19.382	198.23	958.8	1157.0	0.3387	1.3901	1.7238	230
240	24.969	0.01692	16.208	16.223	208.34	952.2	1160.5	0.3531	1.3609	1.7140	240
250	29.823	0.01700	13.804	13.821	216.48	945.5	1164.0	0.3875	1.3323	1.6998	250
260	35.429	0.01709	11.746	11.763	225.64	938.7	1167.3	0.3817	1.3043	1.6860	260
270	41.858	0.01717	10.044	10.061	238.84	931.3	1170.6	0.3958	1.2769	1.6727	270
280	49.203	0.01726	8.628	8.645	249.06	924.7	1173.8	0.4098	1.2301	1.6597	280
290	37.558	0.01735	7.444	7.481	259.31	917.5	1176.8	0.4234	1.2238	1.6472	290
300	67.013	0.01745	6.449	6.468	289.59	910.1	1179.7	0.4399	1.1980	1.8330	300
310	77.68	0.01755	5.609	5.523	279.92	902.6	1182.5	0.4504	1.1727	1.6231	310
320	89.68	0.01765	4.896	4.914	290.23	894.9	1185.2	0.4637	1.1478	1.6115	320
330	103.06	0.01778	4.289	4.307	300.68	887.0	1187.7	0.4760	1.1233	1.6002	330
340	118.01	0.01787	3.770	3.788	311.13	879.0	1190.1	0.4900	1.0992	1.5891	340
350	134.63	0.01799	3.324	3.342	321.53	870.7	1192.3	0.5029	1.0754	1.5783	350
360	153.04	0.01811	2.939	2.957	332.18	862.2	1194.4	0.5158	1.0519	1.3677	350
370	173.37	0.01823	2.606	2.625	342.79	853.5	1196.3	0.5288	1.0287	1.5573	370
380	195.77	0.01826	2.317	2.335	353.45	844.8	1198.1	0.5413	1.0059	1.5471	380
390	220.37	0.01850	2.0651	2.0838	364.17	835.4	1199.6	0.3539	0.9832	1.5371	390

400	247.31	0.01864	1.3447	1.8633	374.97	328.0	1201.0	0.5684	0.9608	1.5272	400
410	278.75	0.01878	1.6512	1.6700	385.53	816.3	1202.1	0.5788	0.9388	1.5174	410
420	308.83	0.01894	1.4811	1.5000	396.77	806.3	1203.1	0.5912	0.9168	1.5078	420
430	343.72	0.01910	1.3308	1.3499	407.79	796.0	1203.8	0.6035	0.8947	1.4982	430
440	281.59	0.01828	1.1979	1.2171	418.90	785.4	1204.3	0.6158	0.8730	1.4887	440
450	422.8	0.0194	1.0799	1.0993	430	774.5	1204.8	0.6290	0.3513	1.4793	450
460	466.9	0.0196	0.9748	0.9944	441.4	783.2	1204.8	0.6402	0.8298	1.4700	460
470	514.7	0.0198	0.8811	0.9009	452.5	751.3	1204.3	0.6523	0.3083	1.4806	470
480	366.1	0.0200	0.7972	0.8172	464.4	739.4	1203.7	0.6645	0.7868	1.4513	480
490	621.4	0.0202	0.7221	0.7423	476.0	726.8	1202.8	0.6768	0.7653	1.4419	490
500	680.8	0.0204	0.6545	0.6749	487.8	713.0	1201.7	0.5887	0.7438	1.4323	500
520	812.4	0.0209	0.5385	0.5594	511.9	688.4	1198.2	0.7130	0.7006	1.4138	520
540	962.5	0.0215	0.4434	0.4549	538.8	656.6	1193.2	0.7374	0.6568	1.2942	540
560	1133.1	0.0221	0.2647	0.3868	582.2	624.2	1186.4	0.7821	0.6121	1.3742	560
580	1325.2	0.0228	0.2988	0.2217	588.9	588.4	1177.3	0.7872	0.5659	1.2532	580
600	1542.9	0.0228	0.2432	0.2668	617.0	548.5	1165.3	0.8131	0.5178	1.3307	600
620	1738.5	0.0247	0.1958	0.2201	646.7	503.8	1150.3	0.8398	0.4664	1.2062	620
640	2059.7	0.0250	0.1538	0.1798	678.6	452.0	1130.3	0.8579	0.4110	1.2789	640
660	2265.4	0.0275	0.1165	0.1442	714.2	390.2	1104.4	0.8987	0.3485	1.2472	660
680	2708.1	0.0305	0.0510	0.1115	757.3	309.9	1067.2	0.9351	0.2219	1.2071	680
700	2093.7	0.0369	0.0392	0.0761	823.3	172.1	995.4	0.9905	0.1484	1.1389	700
705.4	3206.2	0.0503	0	0.0503	902.7	0	902.7	1.0580	0	1.0580	705.4

Density of water = $1 \div$ specific volume.

Latent heat = h_{fg} .

Table 3.7 “*K*” Value for Fluid Flow Perpendicular to a Bank of Staggered Tubes “*N*” Rows Deep

<i>N</i>	<i>K</i>
1	0.24
2	0.27
3	0.29
4	0.30
5	0.31
6	0.32
10	0.33



3.10 TUBE METAL RESISTANCE

The tube metal resistance can be determined from the following equation:

$$r_w = \frac{\Delta X_w}{K_w} = \frac{L}{K} \quad (3.24)$$

where ΔX_w = wall thickness, ft; K_w = thermal conductivity of pipe, Btu/h ft °F (Table 3.10).

Tables 3.8 and 3.9 have basic tube and coil properties for use in Equation (3.10), and Table 3.10 lists the thermal conductivity of different metals at 200 °F.



3.11 APPROXIMATE OVERALL HEAT TRANSFER COEFFICIENTS

Calculation of “*U*” using the previous equation is tedious. Specialists use computer programs to calculate this value. “Quick Look” tables from the GPSA Engineering Data Book (Tables 3.11 and 3.12) provide approximate values of “*U*” for shell-and-tube heat exchangers. For example:

- Exchanging water with 100 psi gas gives a low *U* value, thus requiring a high surface area
- Exchanging water with 1000 psi gas gives a much higher *U* value, thus requiring less surface area in the exchanger
- Exchanging water with water achieves a very high *U* value

Values in the table do not differentiate between tube-side and shell-side fluids. Placement of fluid does make a difference to the *U* value. “Quick Look” graphs from Smith Industries (see Figure 3.30), provide approximate

Table 3.8 Characteristics of Tubing

Tube OD in.	B.W.G Gauge	Thickness in.	Internal Area in.	FT External Surface Per ft Length	FT Internal Surface Per ft Length
1	14	0.083	0.5463	0.2618	0.2183
1	15	0.072	0.5755	0.2618	0.2241
1	16	0.065	0.5945	0.2618	0.2278
1	18	0.049	0.6390	0.2618	0.2361
1	20	0.035	0.6793	0.2618	0.2435
1 1/4	7	0.180	0.6221	0.3272	0.2330
1 1/4	8	0.165	0.6648	0.3272	0.2409
1 1/4	10	0.134	0.7574	0.3272	0.2571
1 1/4	11	0.120	0.8012	0.3272	0.2644
1 1/4	12	0.109	0.8365	0.3272	0.2702
1 1/4	13	0.095	0.8825	0.3272	0.2775
1 1/4	14	0.083	0.9229	0.3272	0.2838
1 1/4	16	0.065	0.9852	0.3272	0.2932
1 1/4	18	0.049	1.042	0.3272	0.3016
1 1/4	20	0.035	1.094	0.3272	0.3089
1 1/2	10	0.134	1.192	0.3927	0.3225
1 1/2	12	0.109	1.291	0.3927	0.3356
1 1/2	14	0.083	1.398	0.3927	0.3492
1 1/2	16	0.065	1.474	0.3927	0.3587
2	11	0.120	2.433	0.5236	0.4608
2	12	0.109	2.494	0.5236	0.4665
2	13	0.095	2.573	0.5236	0.4739
2	14	0.083	2.642	0.5236	0.4801

Table 3.9 Pipe Coil Data

Nom. Size In.	Sch. No.	OD In.	ID In.	Internal Surface Area (ft ² /ft)	External Surface Area (ft ² /ft)
1	S40	1.315	1.049	0.275	0.344
	X80		0.957	0.251	
	160		0.815	0.213	
	XX		0.599	0.157	
2	S40	2.375	2.067	0.541	0.622
	X80		1.939	0.508	
	160		1.687	0.442	
	XX		1.503	0.394	
2 1/2	XXX	2.875	1.375	0.360	0.753
3	S40	3.50	3.068	0.803	0.916
	X80		2.900	0.759	
	160		2.624	0.687	
	XX		2.300	0.602	
4	S40	4.50	4.026	1.054	1.19
	X80		3.826	1.002	
	160		3.438	0.900	
	XX		3.152	0.825	

Table 3.10 Thermal Conductivity of Metals at 200 °F

Material	Conductivity Btu/(h ft °F)
Copper	223
Brass (admiralty)	70
Silicon bronze	15
Stainless steel (18cr-8ni)	8
Inconel	8
90-10 CuNi	30
70-30	18
Monel	15
Titanium	10

Table 3.11 Typical Bare Tube Overall Heat Transfer Coefficients, for Shell-and-Tube Heat Exchangers (Btu/h ft² °F)

Service	<i>U</i>	Service	<i>U</i>
Water with 100 psi	35-40	Water condensers with C ₃ , C ₄	125-135
Gas water with 300 psi	40-50	Water condensers with naphtha	70-80
Gas water with 700 psi	60-70	Water condensers with still ovhd	70-80
Gas water with 1000 psi	80-100	Water condensers with amine	100-110
Gas water with kerosene	80-90	Reboilers w/steam	140-160
Water with MEA	130-150	Reboilers w/hot oil	90-120
Water with air	20-25	100 psi gas w/500 psi gas	50-70
Water with water	180-200	1000 psi gas w/1000 gas	60-80
Oil with oil	80-100	1000 psi gas chiller (gas-C ₃)	60-80
C ₃ with C ₃ liquid	110-130	MEA exchanger	120-130

Notes: Maximum boiling film transfer coefficient. Hydrocarbons: 300-500 Btu/h ft² °F. Water: 200-0 Btu/h ft² °F.

values of “*U*” for exchange from water bath to a natural gas stream in a coil. Figure 3.31 is a nomograph for crude oil streams heated with a water bath.



3.12 DETERMINATION OF PROCESS HEAT DUTY

3.12.1 Overview

The heat required to be added or removed from the process fluids to create the required change in temperature can be in the form of sensible heat, latent heat, or latent heat of vaporization.

Table 3.12 Typical “U” Value Ranges for Shell-and-Tube Exchangers

Service	U, Btu/(h ft ² °F)
---------	-------------------------------

Water coolers	
Gas (to 500 psi)	35-50
Gas (500-1000 psi)	50-80
Gas (over 1000 psi)	80-100
Natural gasoline	70-90
MEA	130-150
Air	15-25
Water	170-200
Water condensers	
Amine regenerator	100-110
Fractionator overhead	70-80
Light hydrocarbons	85-135
Reboilers	
Steam	140-160
Hot oil	90-120
Glycol	10-20
Amine	100-120
General	
Oil-oil	80-100
Propane-propane	100-130
Rich MEA-lean MEA	120-130
Gas-gas (to 500 psi)	50-70
Gas-gas (about 10,000 psi)	55-75
Gas-propane chiller	60-90

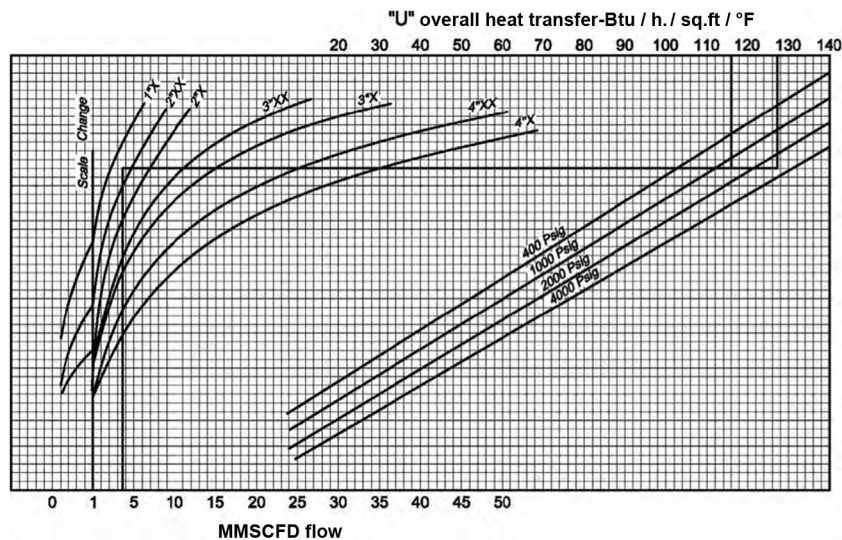


Figure 3.30 Values of U for exchange from water bath to natural gas stream in a coil.

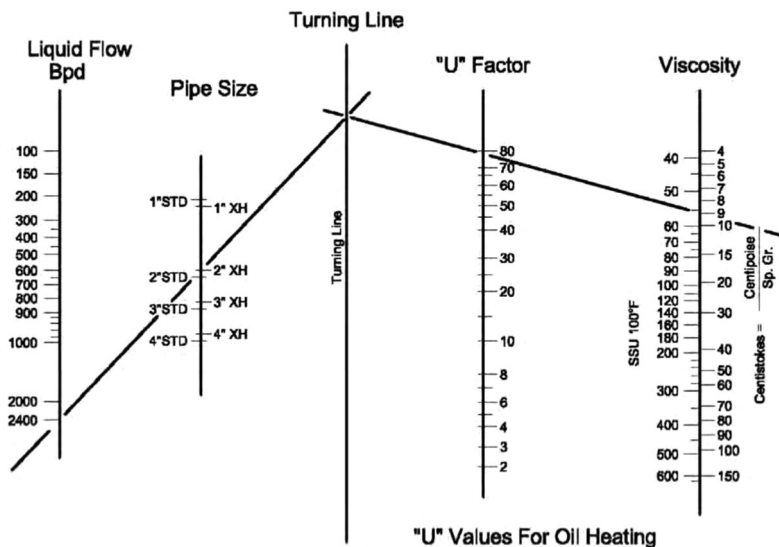


Figure 3.31 Values of U for crude oil streams with water bath heaters.

3.12.2 Sensible Heat

Sensible heat is the amount of heat absorbed or lost by a substance that causes a change in the temperature of the substance. For example, as heat is added to steel, the temperature of the steel increases and can be measured. The general equation is expressed as

$$q_{sh} = WC_p \Delta T \quad (3.25)$$

where q_{sh} = sensible heat duty, Btu/h (kcal/h); W = mass flow rate, lb/h (kg/h); C_p = specific heat of the fluid, Btu/lb-°F (kcal/kg-°C), = Figures 3.32 and 3.33; ΔT = temperature change for a stream °F (°C).

3.12.3 Latent Heat

When a substance changes phases, the heat absorbed or lost is in the form of latent heat. This heat energy cannot be sensed by measuring the temperature. The general equation is expressed as

$$q_{lh} = W\lambda \quad (3.26)$$

where q_{lh} = latent heat duty, Btu/h; W = mass flow rate, lb/h; λ = latent heat, Btu/lb.

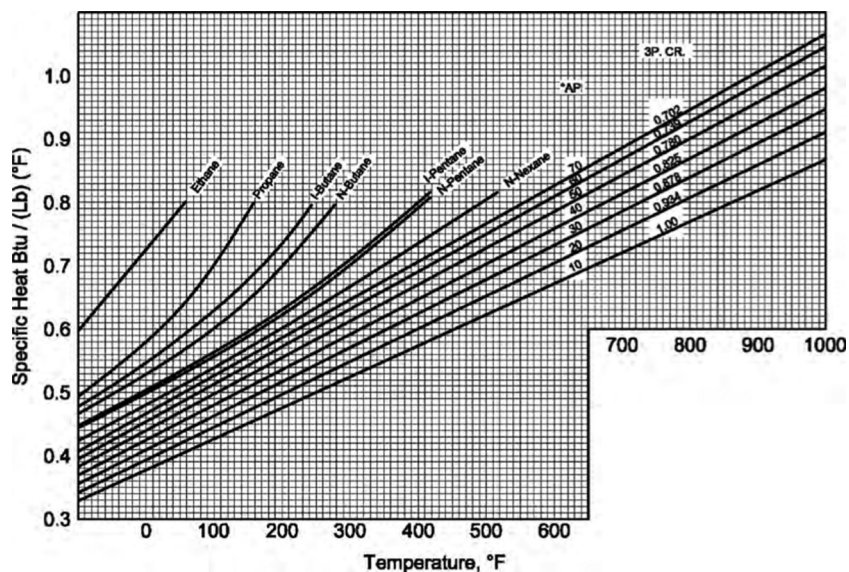


Figure 3.32 Specific heats of hydrocarbon liquids.

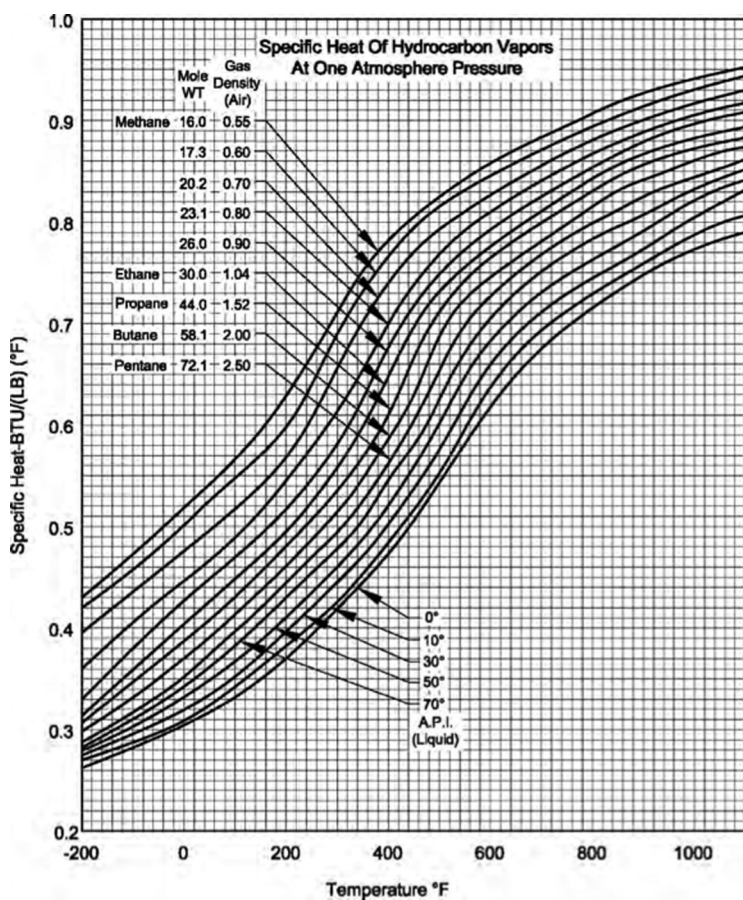


Figure 3.33 Specific heats of hydrocarbon vapors.

Table 3.13 Latent Heat of Vaporization for Hydrocarbon Compounds

Compound	Heat of Vaporization 14.696 psia at Boiling Point, Btu/lb
Methane	219.22
Ethane	210.41
Propane	183.05
<i>n</i> -Butane	165.65
Isobutane	157.53
<i>n</i> -Pentane	153.59
Isopentane	147.13
Hexane	143.95
Heptane	136.01
Octane	129.53
Noane	123.76
Decane	118.68

Latent heat of vaporization for hydrocarbon compounds is given in Table 3.13. Latent heat of vaporization of water is given by h_{fg} in the steam tables.

3.12.4 Heat Duty for Multiphase Streams

The process heat duty when more than one phase exists in the process stream can be calculated using the following equation:

$$q_p = q_g + q_o + q_w \quad (3.27)$$

where q_p = overall heat duty, Btu/h; q_g = gas heat duty, Btu/h; q_o = oil heat duty, Btu/h; q_w = water heat duty, Btu/h.

3.12.5 Natural Gas Sensible Heat Duty at Constant Pressure

The sensible heat duty for natural gas at constant pressure is expressed as

$$q_g = 41.7(T_2 - T_1)C_gQ_g \quad (3.28)$$

where T_2 = inlet temperature, °F; T_1 = outlet temperature, °F; C_g = gas heat capacity, Btu/Mscf °F; Q_g = gas flow rate, MMscfd.

Gas heat capacity, C_g , is determined at atmospheric conditions and then corrected for temperature and pressure based on reduced pressure and temperature.

3.12.6 Gas Heat Capacity

The general equation for gas heat capacity is expressed as

$$C_g = 2.64 [29.SC_p + \Delta C_p] \quad (3.29)$$

where C_g =gas-specific heat at one atmosphere pressure, Btu/lb °F; ΔC_p =correction factor (from Figure 3.34); S =gas-specific gravity.

The correction factor, ΔC_p , is obtained from Figure 3.34 when the values of P_r and T_r are known.

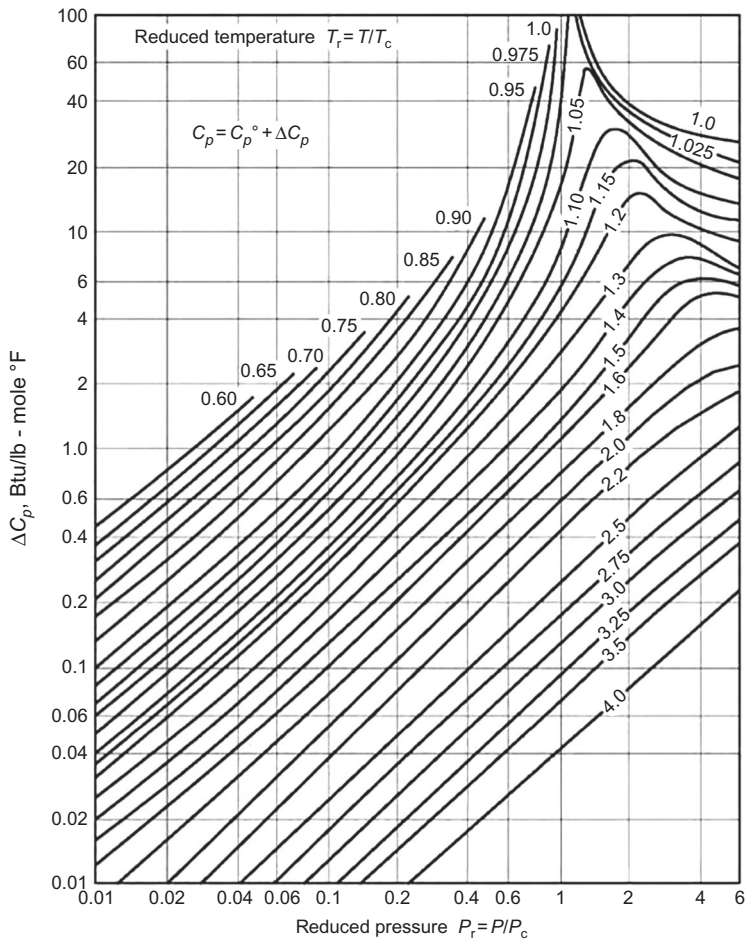


Figure 3.34 Heat capacity correction factor.

$$T_r = \frac{T_a}{T_c} \quad (3.30)$$

$$P_r = \frac{P}{P_c} \quad (3.31)$$

where P_r = gas reduced pressure; P = gas pressure, psia; P_c = gas pseudo critical pressure, psia, = [Figure 3.35](#); T_r = gas reduced temperature; T_a = gas average temperature, °R, = $\frac{1}{2}(T_1 + T_2)$; T_c = gas pseudo critical temperature, °R, = [Figure 3.35](#).

3.12.7 Calculation of Gas Pseudo Critical Pressure and Temperature

The gas pseudo critical pressure and temperature can be approximated from [Figure 3.35](#). It is calculated as a weighted average of critical temperature and pressure of various components on a mole fraction basis (refer to [Table 3.14](#)

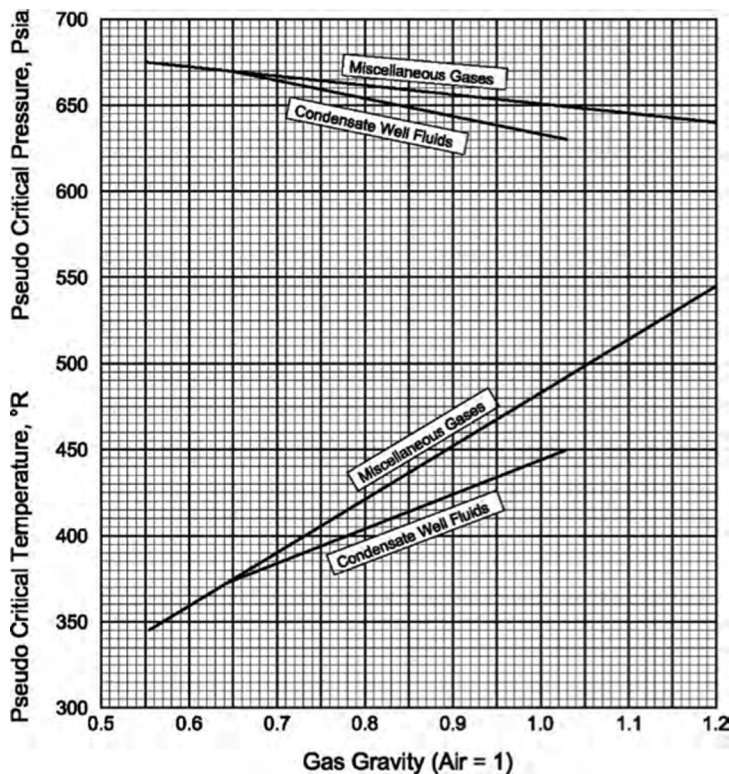


Figure 3.35 Pseudo critical properties of natural gas.

Table 3.14 Estimate of Specific Gravity; Pseudo Critical Temperature and Pseudo Critical Pressure, for Example, Field

	A	B	C	D
	Mole Percent			
	Gas	Molecular		
Component	Composition	Weight	Critical Temp °R	Critical Psia
CO ₂	4.03	44.010	547.87	1071.0
N ₂	1.44	28.013	227.3	493.0
H ₂ S	0.0019	34.076	672.6	1036.0
C ₁	85.55	16.043	343.37	667.8
C ₂	5.74	30.070	550.09	707.0
C ₃	1.79	44.097	666.01	616.3
iC ₄	0.41	58.124	734.98	529.1
nC ₄	0.41	58.124	765.65	550.7
iC ₅	0.20	72.151	829.10	490.4
nC ₅	0.13	72.151	845.70	488.6
C ₆	0.15	86.178	913.70	436.9
C ₇₊	0.15	147.0	1112.0	304.0
Computed value	100.00	19.48	374.55	680.33
Computation	$\sum A_i$	$\sum [(A_i B_i) / \sum A_i]$	$\sum [(A_i C_i) / \sum A_i]$	$\sum [(A_i D_i) / \sum A_i]$
Specific gravity	=19.48/29			
	=0.67			

for sample calculations). For greater precision, a correction for H₂S and CO₂ content should be made.

3.12.8 Oil Sensible Heat Duty

The sensible heat duty for the oil phase is:

$$q_o = 14.6(\text{SG})(T_2 - T_1)C_o Q_o \quad (3.32)$$

where C_o = oil-specific heat, Btu/lb °F (Figure 3.35); Q_o = oil flow rate, BPD; SG = oil-specific gravity (water = 1).

3.12.9 Water Sensible Heat Duty

The duty for heating free water may be determined from the following equation assuming a water heat capacity of 1.0 Btu/lb °F.

$$q_w = 14.6(T_2 - T_1)Q_w \quad (3.33)$$

where Q_w = water flow rate, bwpd.

3.12.10 Heat Duty Where There Are Phase Changes

The heat duty where there are phase changes is best determined by performing flash calculations so as to determine the heat loss or gain by the change in enthalpy. For a “quick look” hand approximation of the total heat duty, calculate the sensible heat for both the gas and the liquid phases of each component and then sum all latent and sensible heats.

3.12.11 Heat Loss to Atmosphere

The heat loss to the atmosphere can be calculated using the general equation:

$$q = UA\Delta t_m \quad (3.34)$$

“ U ” can be calculated from a modification of Equation (3.18) assuming the inside film coefficient is very large compared to the outside film coefficient.

Under this condition, Equation (3.18) reduces to:

$$\frac{1}{U} = \frac{1}{h_o} + \frac{\Delta X_1}{K_1} + \frac{\Delta X_2}{K_2} \quad (3.35)$$

where $h_o = 1 + 0.22 V_w$ ($V_w < 16$ ft/s), $= 0.53 V_w^{0.8f}$ ($V_w > 16$ ft/s); V_w = wind velocity (ft/s), $= 1.47 \times$ (mph); ΔX_1 = shell thickness, ft; K_1 = shell thermal conductivity, Btu/h ft °F, $= 30$ for carbon steel (Table 3.11); ΔX_2 = insulation thickness, ft; K_2 = insulation thermal conductivity, Btu/h ft °F, $= 0.03$ for mineral wool.

For preliminary calculations we often assume:

- 5-10% of q for uninsulated equipment
- 1-2% of q for insulated vessels

3.12.12 Heat Transfer from a Fire Tube

A fire tube contains a flame burning inside a piece of pipe, which is in turn surrounded by the process fluid. Radiant and convective heat transfer occurs from the flame to the inside surface of the fire tube. Conductive heat transfer occurs through the wall thickness of the tube and convective heat transfer from the outside surface of that tube to the fluid being heated. It is difficult to solve for the heat transfer in terms of an overall heat transfer coefficient. Rather, what is done is to size the fire tube by using a heat flux rate. The heat flux rate represents the amount of heat that can be transferred from the fire tube to the process per unit area of outside surface of the fire tube. Common heat flux rates are given in Table 3.15. The required fire tube area is given by

Table 3.15 Common Heat Flux Rates	
Medium Being Heated	Design Heat Flux Rate Btu/h ft ²
Water	10,000
Boiling water	10,000
Crude oil	8000
Heating medium oils	8000
Glycol	7500
Amine	7500

$$(\text{Surface Area of Fire Tube}) = \frac{\text{Heat Duty Included Losses (Btu/h)}}{\text{Design Flux Rate (Btu/h - ft}^2\text{)}} \tag{3.36}$$

Example 3.3: Determination of Total Heat Duty from a Fire Tube Given

The total heat duty (sensible heat, latent heat duty, heat losses to atmosphere) is 1 MMBtu/h and water is being heated with a heat flux rate of 10,000 Btu/h ft².

Determine

The required surface area of the fire tube.

Solution

Substituting into Equation (3.36)

$$\begin{aligned} \text{Surface Area of Fire Tube} &= (1,000,000 \text{ Btu/h}) / (10,000 \text{ Btu/h ft}^2) \\ &= 100 \text{ ft}^2 \text{ of fire tube area required} \end{aligned}$$

3.12.13 Natural Draft Fire Tubes

Minimum cross-sectional area of the fire tube is set by limiting the heat release density to 21,000 Btu/h in.²

At heat release densities above this value, the flame may become unstable because of insufficient air. Using this limit, a minimum fire tube diameter is established by

$$d^2 = \frac{\text{Burner Heat Release Density (Btu/h)}}{16,500} \tag{3.37}$$

where *d* = minimum fire tube diameter, in.

Note that the burner heat release density will be somewhat higher than the heat duty (including losses) because a standard burner size will be chosen

Table 3.16 Standard Burner Sizes and Minimum Diameter

Btu/h	Minimum Diameter, in.
100,000	2.5
250,000	3.9
500,000	5.5
750,000	6.7
1,000,000	7.8
1,500,000	9.5
2,000,000	11.0
2,500,000	12.3
3,000,000	13.5
3,500,000	14.6
4,000,000	15.6
5,000,000	17.4

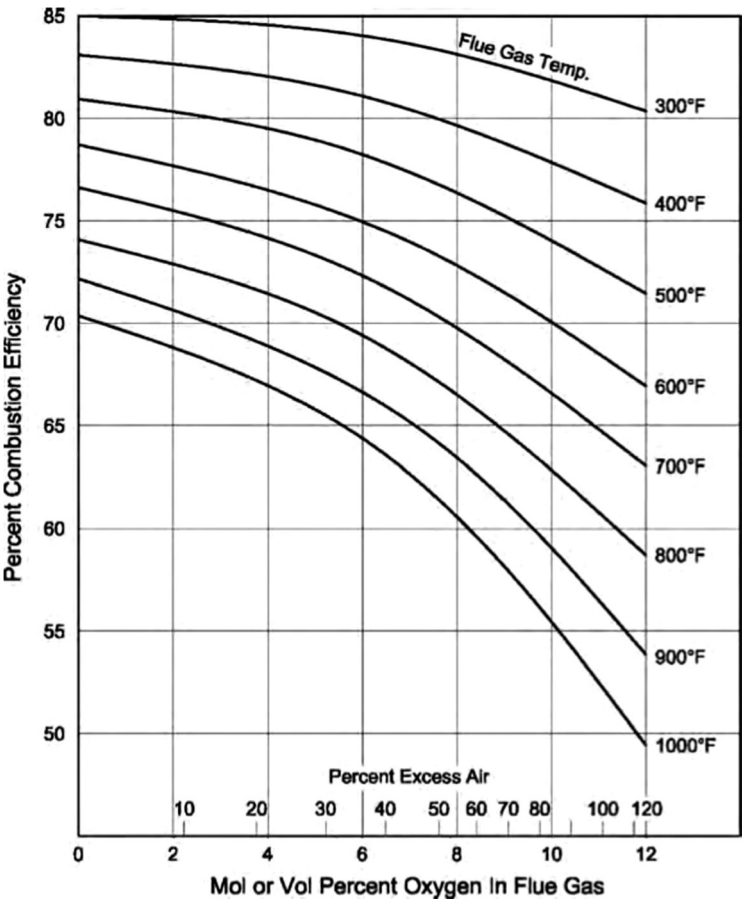


Figure 3.36 Approximate combustion efficiency of natural gas (1050 Btu/scf) in emulsion treaters.

slightly larger than that required. Standard burner sizes and minimum fire tube diameters are included in Table 3.16. Figure 3.36 approximates the combustion efficiency of natural gas (1050 Btu/scf) in emulsion treaters.

3.12.14 Procedure to Size a Shell-and-Tube Heat Exchanger

The procedure to size a shell-and-tube heat exchanger is as follows:

- Calculate LMTD
- Calculate “ U ”

Use heat transfer equations

- Use GPSA procedure in the Engineering Data Book
- Use approximate overall heat transfer coefficient from tables, nomographs, or graph
- Calculate the process heat duty
- Calculate the heat transfer area required

REFERENCES

- [1] F.L. Eisinger, Unusual acoustic vibration of a shell-and-tube process heat exchanger, J. Press. Vessel Technol. 116 (1994) 141–149.
- [2] F.L. Eisinger, R.E. Sullivan, Experience with unusual acoustic vibration in heat exchanger and steam generator tube banks, J. Fluids Struct. 10 (1996) 99–107.



Heat Exchanger Configurations



4.1 OVERVIEW

Heat exchangers used in oil and gas facilities are configured as follows:

Fluid-fluid

- Shell and tube
- Double pipe
- Plate and frame

Cooling with air

- Air-cooled exchangers
- Cooling towers

Bath heaters

- Direct
- Indirect

The remainder of this chapter discusses basic concepts in sizing and selecting heat exchangers.



4.2 SHELL-AND-TUBE EXCHANGERS

4.2.1 Tubular Exchanger Manufacturers Association

The Tubular Exchanger Manufacturers Association (TEMA) defines the various types of shell-and-tube exchangers as well as design and construction practices. Classes of interest are:

- Class “C”—Less stringent, used onshore, and with temperatures above -20°F
- Class “R”—Offshore and cold temperature service

4.2.2 Common Services

- Liquid-liquid
- Liquid-vapor
- Vapor-vapor

4.2.3 Components (Figures 4.1 and 4.2)

- Shell with two nozzles
- Tube sheet heads
- Transverse baffles

4.2.4 Configuration Considerations

- Fluids involved
- Corrosion potential
- Problems of cleaning
- Pressure drop
- Heat transfer efficiency
- Heat exchanger selection is not routine
- Tube lengths
 - 20 ft
 - 40 ft

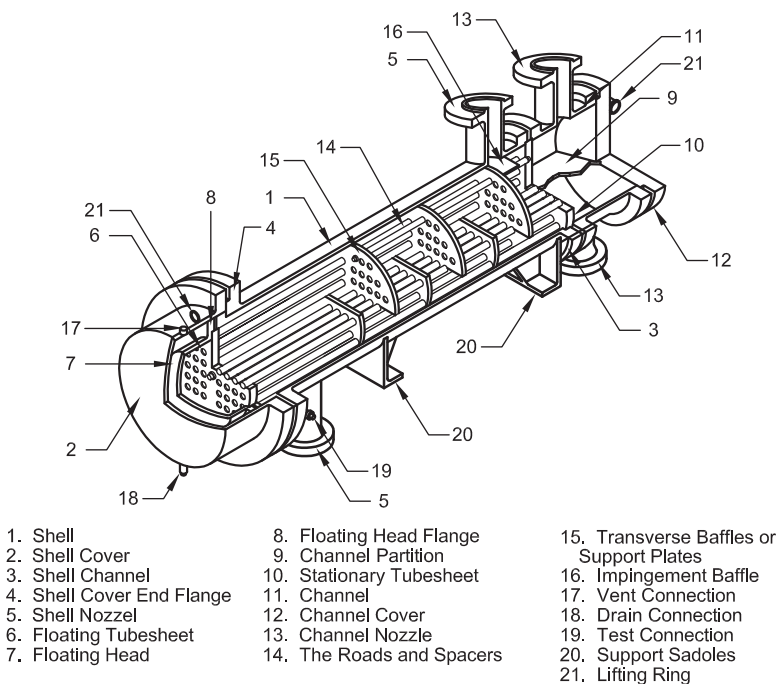


Figure 4.1 Components of shell-and-tube exchangers.

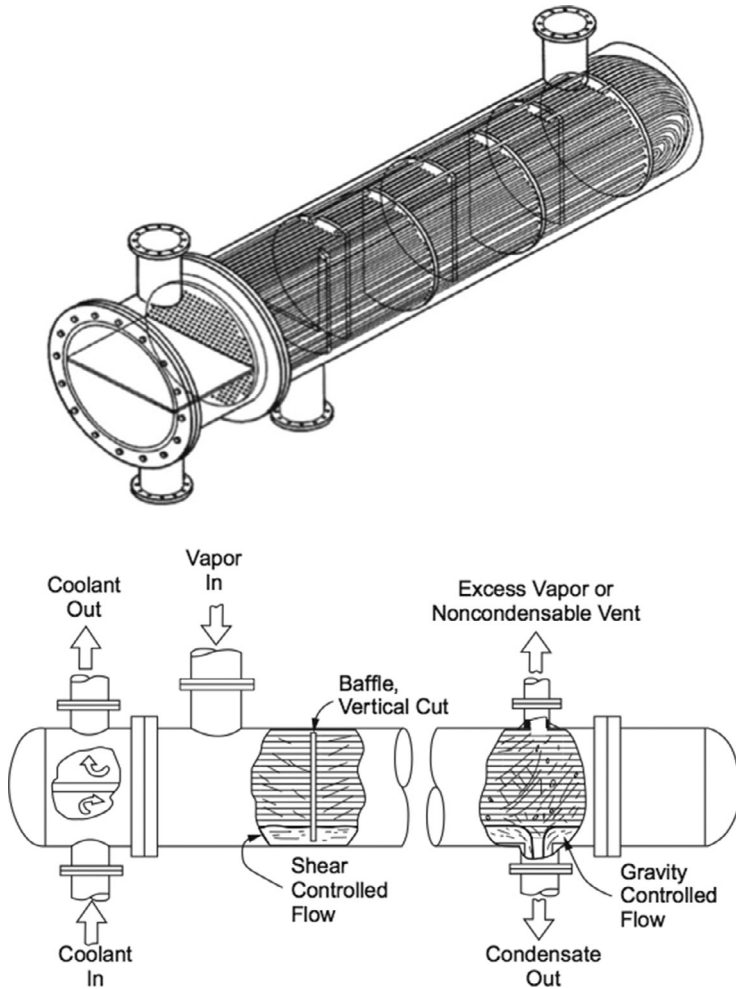


Figure 4.2 Cutaway of two shell-and-tube heat exchangers.

4.2.5 Baffles

Baffles direct the flow of both tube-side and shell-side fluids (Figures 4.3–4.6).

4.2.5.1 Pass Partition

- Forces fluid to flow through several groups of parallel tubes
- Increases the number of passes

4.2.5.2 Two-Pass Fixed Exchanger

- Tube fluid passes through fraction of tubes
- Maintains high tube velocity
- Increases heat transfer

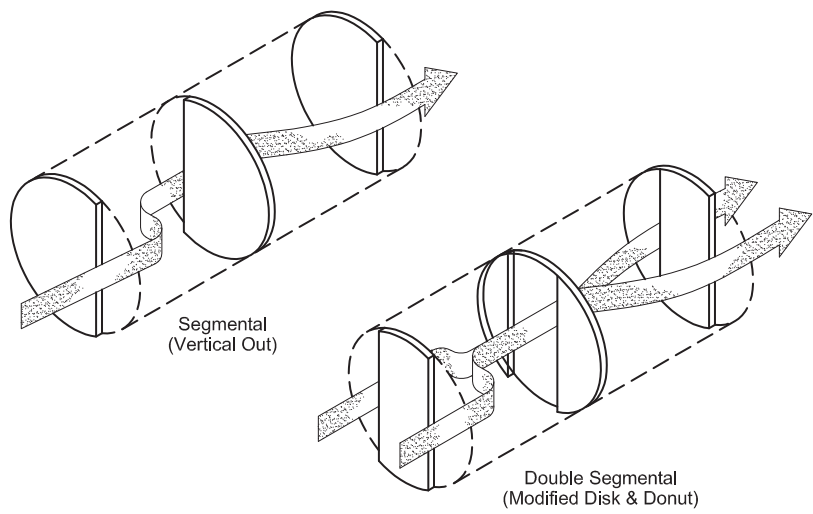


Figure 4.3 Exchanger baffles.

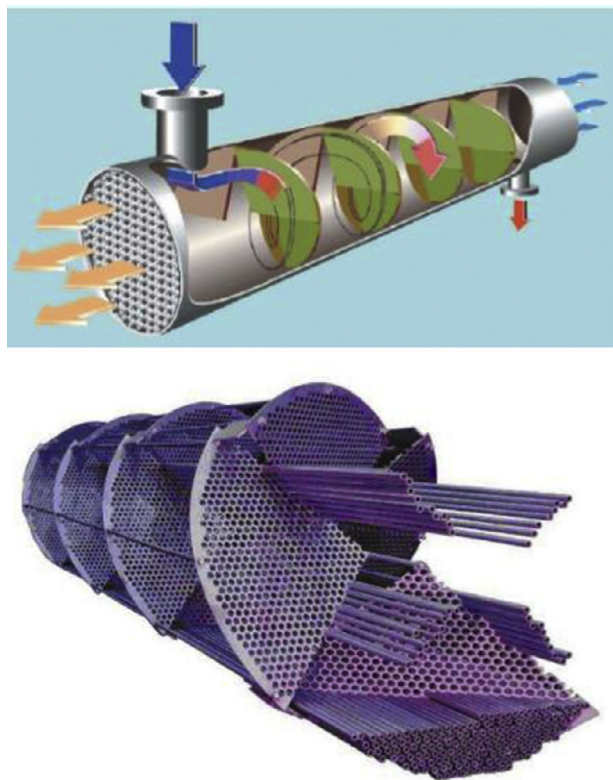


Figure 4.4 Shell-and-tube baffles—helical baffles.



Figure 4.5 Shell-and-tube baffles—helical baffles.

4.2.5.3 Impingement

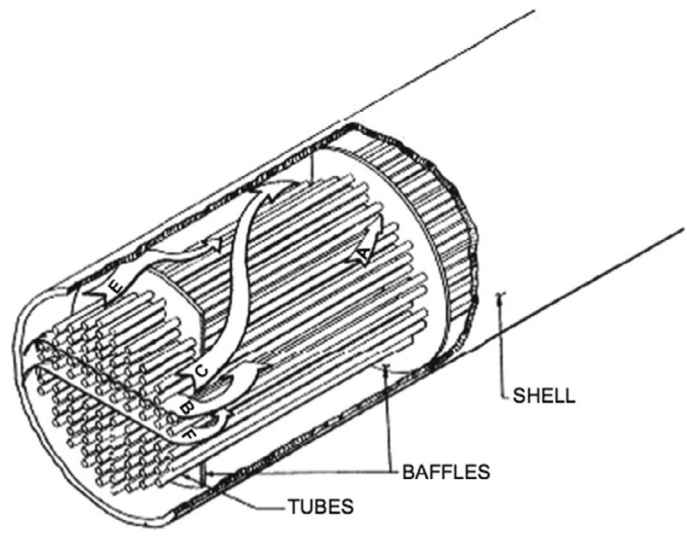
- Is placed opposite shell-side inlet nozzle
- Disperses fluid around tubes
- Prevents impingement and erosion of tubes

4.2.5.4 Transverse or Support (Figures 4.7–4.10)

- Supports the tubes that pass through holes in the baffles
- Maintains turbulence of shell-side fluid
- Results in greater heat transfer

Examples

- 50% (half circle yields least pressure drop but maintains support)
- Usually 75% of height of inside diameter of shell
- Flows up and down
- Flows side to side (used when gas/liquid mixture flows through shell)



Streams

- A – Tube-to-baffle leakage (partly effective)**
- B – Flow across tube bundle (most effective)**
- C – Flow around perimeter of tube bundle**
- D – Baffle-to-shell leakage**
- E – Flow through tube pass partition lanes**

Figure 4.6 Shell-side flow paths.

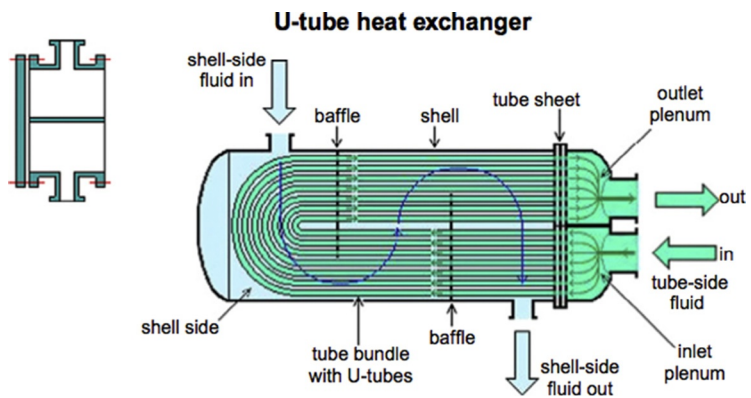


Figure 4.7 Shell-and-tube heat exchanger baffles.

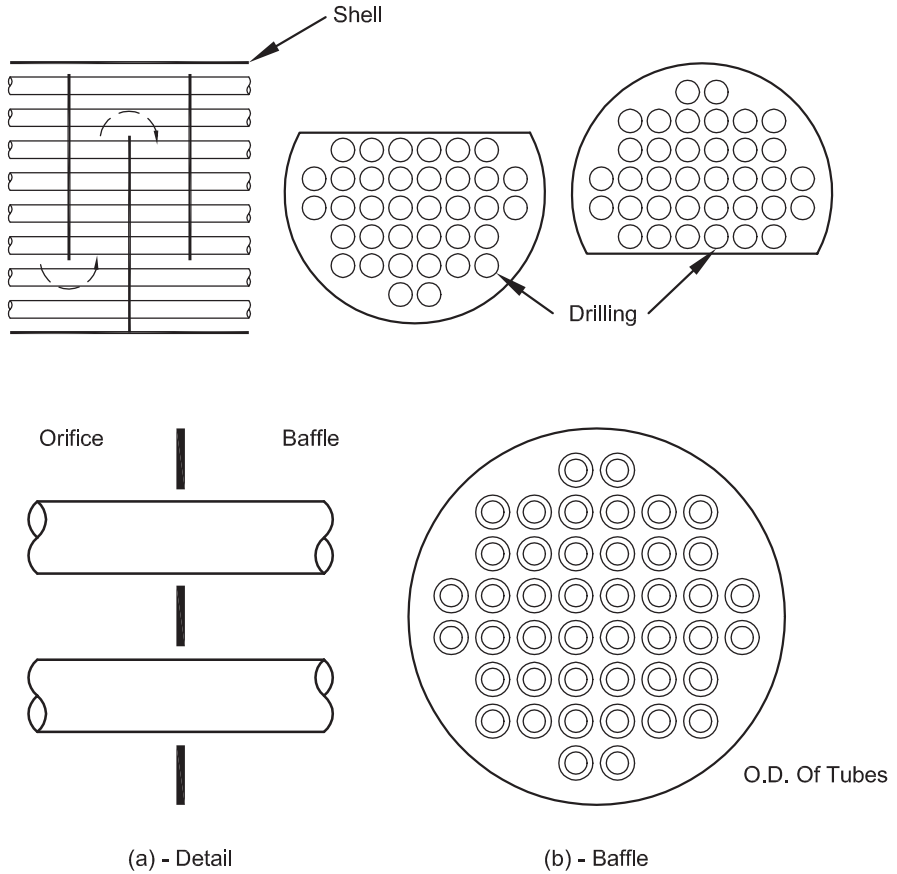


Figure 4.8 Transverse baffle detail.

4.2.5.5 Longitudinal

- Forces shell-side fluid to make more than one pass through the exchanger

4.2.6 Use of Baffles

The main functions of a baffle are to

- Hold tubes in position (preventing sagging) both in production and in operation
- Prevent the effects of vibration, which is increased with both fluid velocity and the length of the exchanger
- Direct shell-side fluid flow along the tube field; this increases fluid velocity and the effective heat transfer coefficient of the exchanger

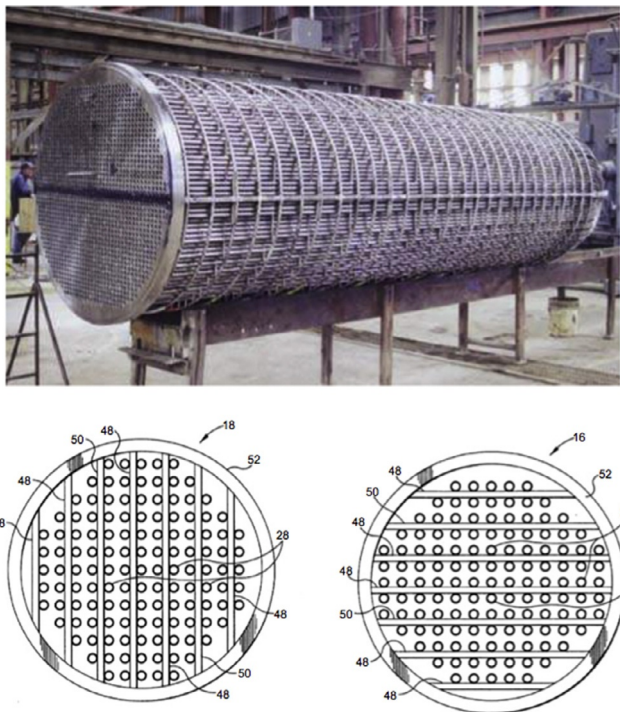


Figure 4.9 Orifice baffle.

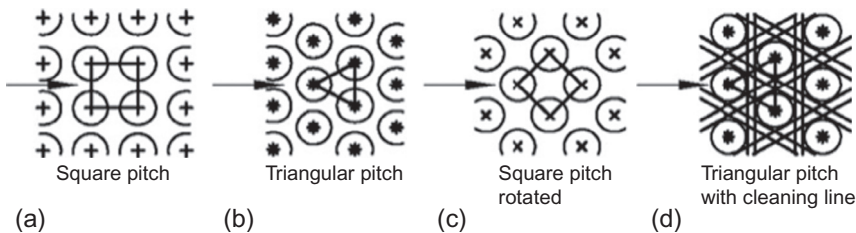


Figure 4.10 Shell-and-tube baffles—rod baffles.

4.2.7 Types of Baffles

Implementation of baffles are decided on the basis of

- Size
- Cost
- Ability to lend support to tube bundles' direct flow.

Often, this is linked to available pressure drop, size, and number of passes within the exchanger. Special allowances/changes are also made for finned tubes. Different types of baffles include

- Segmental baffles (of which single-segment is the most common)
- Rod or bar baffles (give a uniform shell-side flow)
- Helical baffles (similar to segmental types with less pressure drop for the same size exchanger)
- Longitudinal flow baffles (used in a two-pass shell)
- Impingement baffle plates (used to protect bundle when entrance velocity is high)

4.2.8 Installation of Baffles

Baffles deal with the concern of support tubes and fluid direction in heat exchangers. It is vital that they are spaced correctly at installation. The minimum baffle spacing is the greater of 50 mm or one-fifth the inner shell diameter. The maximum baffle spacing depends on the material and size of tubes. Guidelines for this are established by TEMA. There are also segments with a “no tubes in window” design that affects the acceptable spacing within the design.

An important design consideration is that no recirculation zones or dead spots form—both of which are counterproductive to effective heat transfer.

4.2.9 Tubes

Heat exchanger tubes are not to be confused with steel pipe (extruded). The outside diameter (OD) is the actual OD in inches (strict tolerance). Common tube sizes include:

- 5/8 in.
- 3/4 in.
- 1 in.

Tubes are available in a variety of metals such as:

- Steel
- Copper
- Brass
- 70-30 copper nickel
- Bronze
- Aluminum
- Stainless steel

Tube wall thickness is defined by Birmingham Wire Gauge (BWG).

4.2.10 Shells

Less than 24-in. outside diameter (OD)

- Fabricated from pipe

- Nominal diameters are used
- Wall thickness determined as in piping

Greater than 24-in. OD

- Fabricated from rolled steel plate
- Wall thickness determined from ASME Boiler and Pressure Vessel Code
- TEMA Standards specify minimum weight of shell

4.2.11 Tube Pitch ([Figure 4.11](#))

Tube holes cannot be drilled too close together as it weakens the tube sheet. The “pitch” is the shortest center-to-center distance between tubes. The “clearance” is the shortest distance between the outside of two adjacent tubes.

4.2.11.1 *Square Pitch Is Easy to Clean*

- Offers less ΔP when shell-side fluid flows perpendicular to tube axis
- Common pitches:
 - $\frac{3}{4}$ in. OD \times 1 in.
 - 1 in. OD \times $1\frac{1}{4}$ in.

4.2.11.2 *Triangular Pitch Common Pitches*

- $\frac{3}{4}$ in. OD: 15/16 in.
- $\frac{3}{4}$ in. OD: 1 in.
- 1 in. OD: $1\frac{1}{4}$ in.

4.2.12 Options

There are many different arrangements of shells, tubes, and baffles in heat exchangers.

[Figure 4.12](#) shows TEMA nomenclature for shell-and-tube exchangers. The nomenclature assigns letter designations for the stationary head, shell, and rear head. Common combinations are shown in [Figures 4.13–4.15](#).

Three-letter type description:

- First letter designates front end
- Second letter designates shell type
- Third letter designates back end

4.2.13 Classification of Exchangers

Three-element shorthand

- **First element**—nominal diameter
- **Second element**—nominal length
- **Third element**—type

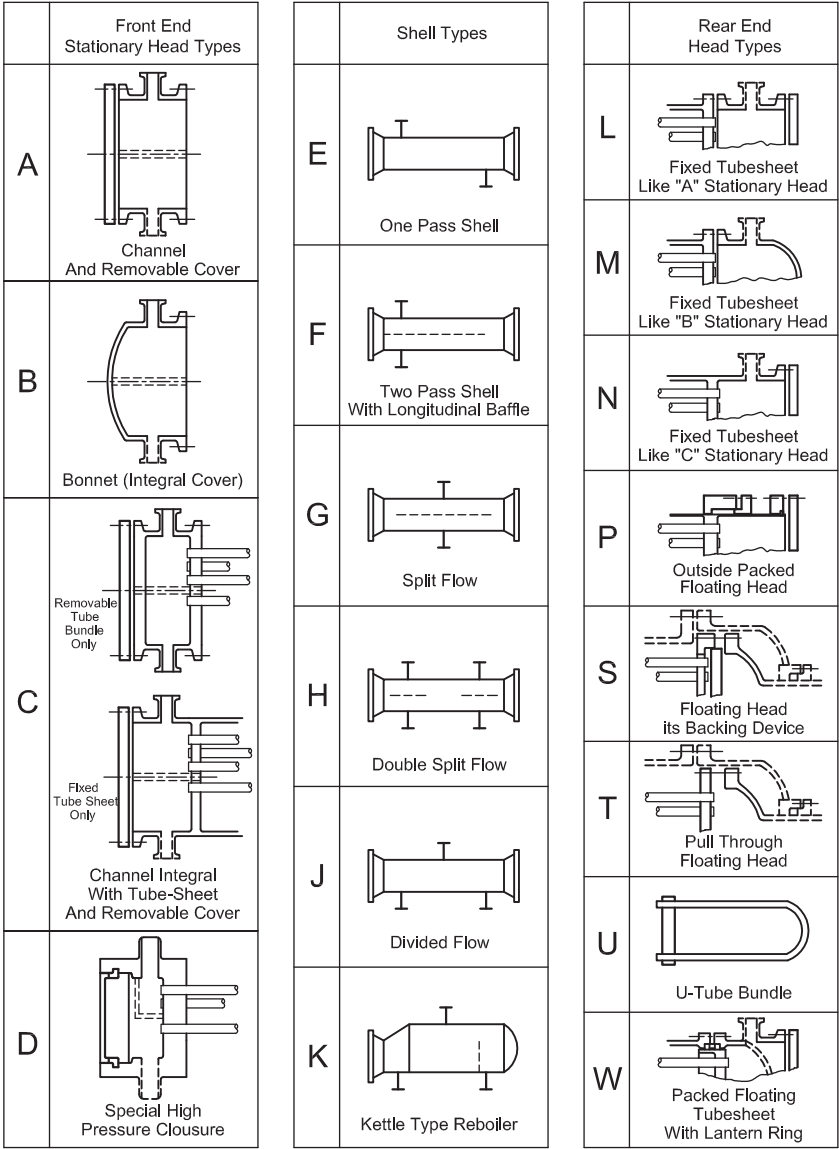
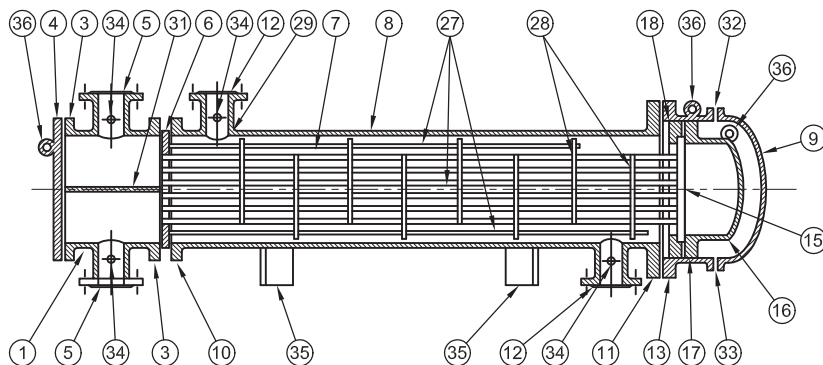


Figure 4.11 Common tube layouts for shell-and-tube heat exchangers.

4.2.13.1 Nominal Diameter

The nominal diameter is the inside diameter of the shell, in inches, rounded off to the nearest integer. For kettle reboilers and chillers (that have a narrow end and a fat end), the nominal diameter is the port (narrow end) followed by the shell diameter, each rounded to the nearest integer.

- | | |
|---|--|
| 1. Stationary Head-Channel | 20. Slip-On Backing Flange |
| 2. Stationary Head-Bonnet | 21. Floating Head Cover-External |
| 3. Stationary Head Flange-Channel Or Bonnet | 22. Floating Tubesheet Skirt |
| 4. Channel Cover | 23. Packing Box |
| 5. Stationary Head Nozzle | 24. Packing |
| 6. Stationary Tubesheet | 25. Packing Gland |
| 7. Tubes | 26. Lantern Ring |
| 8. Shell | 27. Tierods & Spacers |
| 9. Shell Cover | 28. Transverse Baffles Or Support Plates |
| 10. Shell Flange- Stationary Head End | 29. Impingement Plate |
| 11. Shell Flange-Rear Head End | 30. Longitudinal Baffle |
| 12. Shell Nozzle | 31. Pass Partition |
| 13. Shell Cover Flange | 32. Vent Connection |
| 14. Expansion Joint | 33. Drain Connection |
| 15. Floating Tubesheet | 34. Instrument Connection |
| 16. Floating Head Cover | 35. Support Saddle |
| 17. Floating Head Flange | 36. Lifting Lug |
| 18. Floating Head Backing Device | 37. Support Bracket |
| 19. Split Shear Ring | 38. Weir |
| | 39. Liquid Level Connection |



1-Pass Shell, 2-Pass Tube Exchanger

Figure 4.12 TEMA nomenclature for shell-and-tube exchangers.

4.2.13.2 Nominal Length

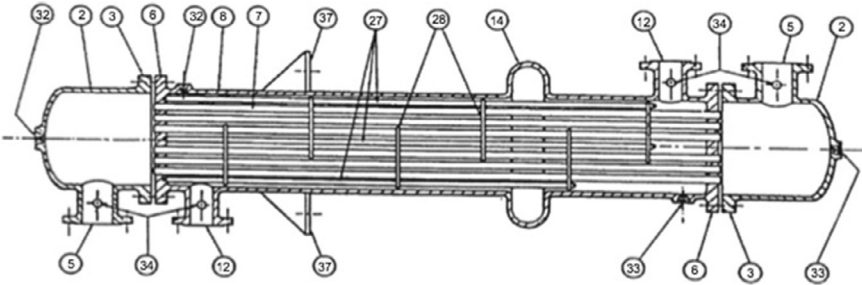
The nominal length is the tube length, in inches. The tube length for straight tubes is taken as the actual overall length. The tube length for U tubes is taken as the straight length from the end of tube to the bend tangent.

4.2.13.3 Type

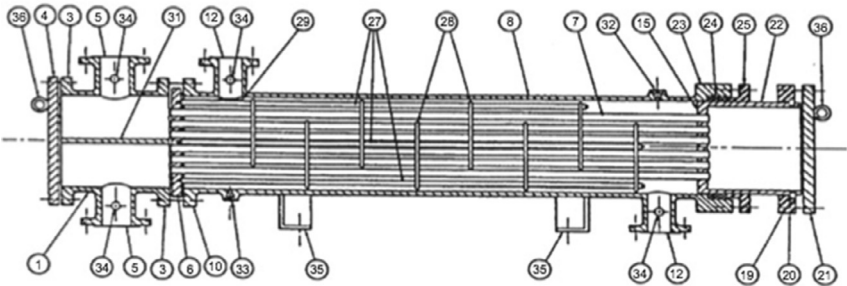
Letters describe the stationary head, shell (omitted for bundles only), and rear head in that order.

4.2.14 Examples of Classification

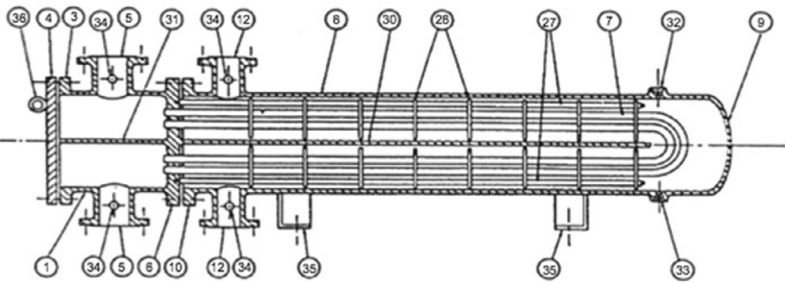
A fixed tube sheet (L) exchanger with removable channel and cover (A), single-pass shell (E), and 23-in. inside diameter with tubes 16 ft long is denoted as Size 23-192 Type AEL.



1-Pass Shell and Tube with Expansion Joint on Shell Side



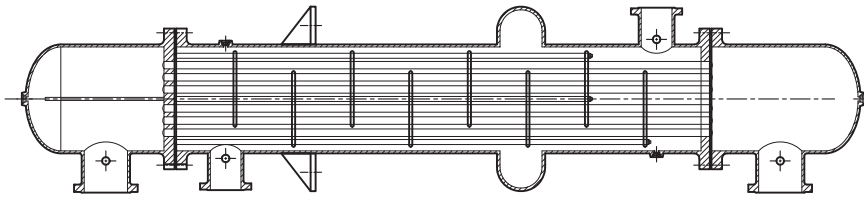
1-Pass Shell and 2-Pass Tube



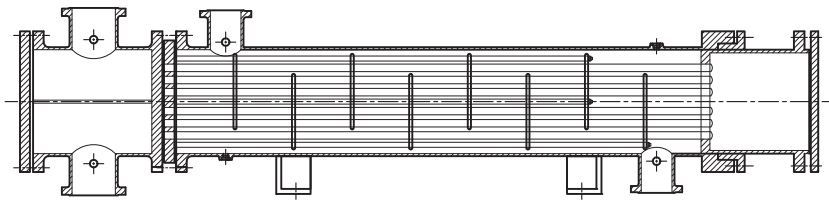
Hairpin

Figure 4.13 Heat exchanger components (type AES).

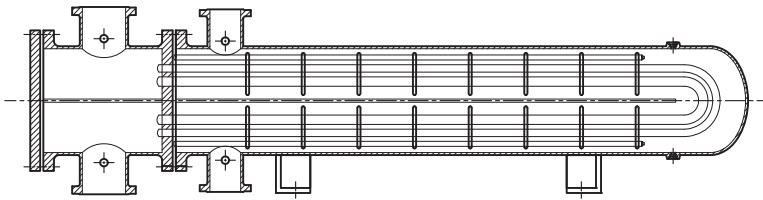
A pull-through floating head (T), kettle type reboiler (K) having stationary head integral with tube sheet (C), 23-in. port diameter and 37-in. inside shell diameter with tubes 16-ft-long is denoted as Size 23/37-192 Type CKT.



1-Pass Shell and Tube with Expansion Joint on Shell side



1-Pass Shell and 2-Pass Tube



Hairpin

Figure 4.14 Examples of tubular exchangers.

4.2.15 Selection of Types

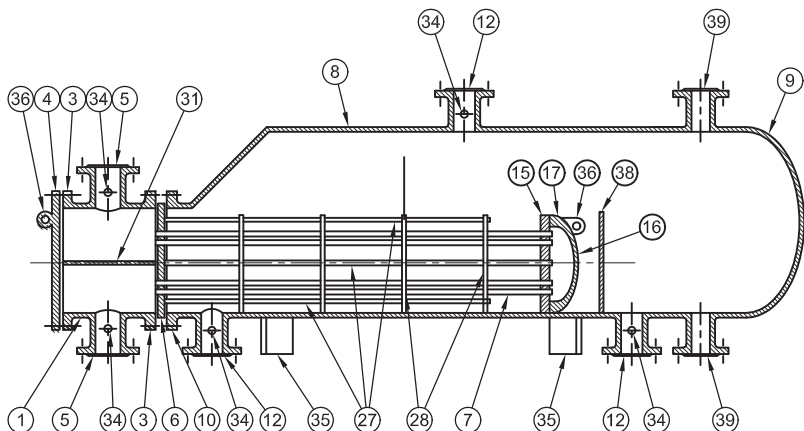
When selecting an exchanger, one must know the advantages and disadvantages. The three types are:

- Fixed tube sheet
- U tube
- Floating head

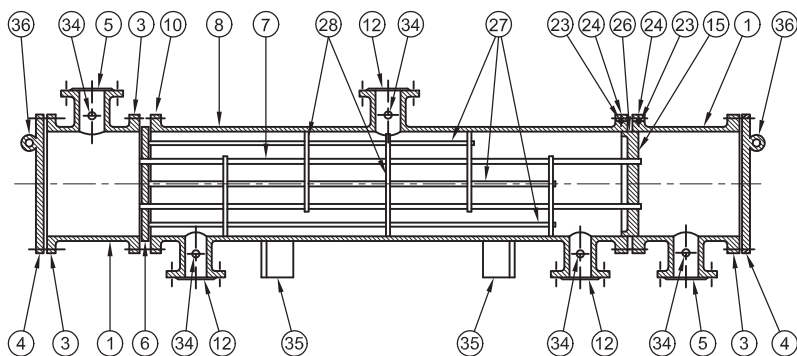
4.2.16 Selecting Heat Exchanger Components

When selecting heat exchanger components one must

- Ensure long-term efficient heat transfer performance
- Simplify maintenance
- Prevent fouling/mechanical problems



Kettle Reboiler



Thermosiphon Reboiler

Figure 4.15 Common types of reboilers.

4.2.17 Picking Heat Exchanger Components

4.2.17.1 Picking Shell Style

- “E type”—Typical shell choice and the least expensive; consider other types if there are pressure drop limitations
- “F type”—Optimum theoretical type, but has problems with longitudinal leakage
- “J type”—Cuts flow path in half and flow rate per pass in half
- “X type”—Has the shortest flow path; used in reboilers with shell-side boiling and for services with very little available pressure drop

U tubes cost the least and are the easiest to maintain. Using a full support plate at the U-bend tangent prevents vibration.

4.2.17.2 Picking Baffles

Baffles can be one of four types (refer to [Figures 4.16](#) and [4.17](#))

- Segmental
- Double segmental
- Triple segmental
- Segmental no tubes in window

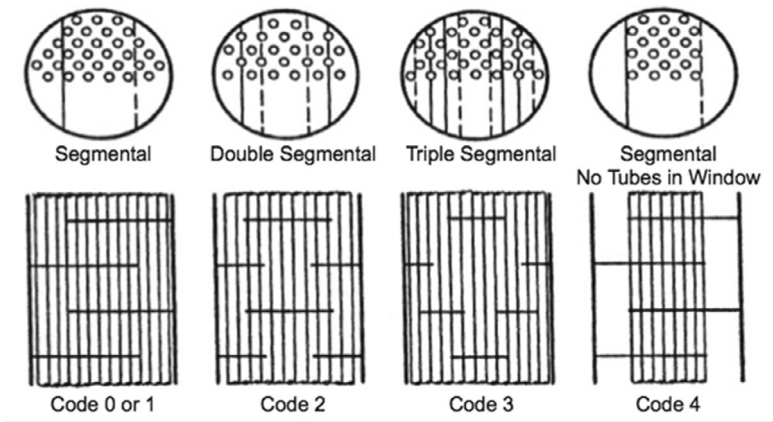


Figure 4.16 Baffle plates.

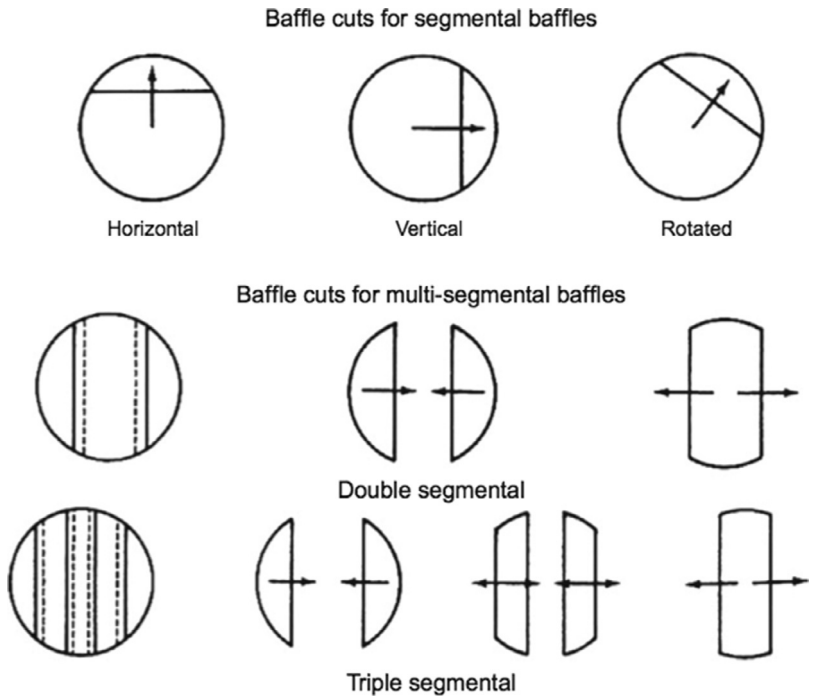


Figure 4.17 Shell-and-tube baffles.

For transverse (central) baffle spacing

- Start with a typical spacing of 20–50% shell inner diameter for sensible heat transfer Provide for economic gradient and good heat transfer
- Meet TEMA requirements for support and structural strength

For baffle cuts

- Single segmental with vertical cuts (for maintenance)
- Most efficient cut is 20–30% shell-side diameter

4.2.17.3 Picking Components

Details are important. For example:

- Use seal bars and dummy tubes to reduce bypassing
- Use support plates to prevent tube vibration
- Include properly placed impingement rods
- Include pull holes for bundle extraction

See also [Figures 4.18–4.26](#).

4.2.18 Fixed Tube Sheet

4.2.18.1 Advantages

- Least expense
- Few gaskets
- Can replace individual tubes
- Provides maximum protection against shell-side leakage

4.2.18.2 Disadvantages

- Cannot clean or inspect shell
- Limited to temperature difference of 200 °F without costly shell expansion joint
- Cannot replace bundle

4.2.18.3 Heads

- A and L are most common
- B and M are for large diameter and high pressures

4.2.18.4 Uses

- Clean shell fluid
- Low differential temperature

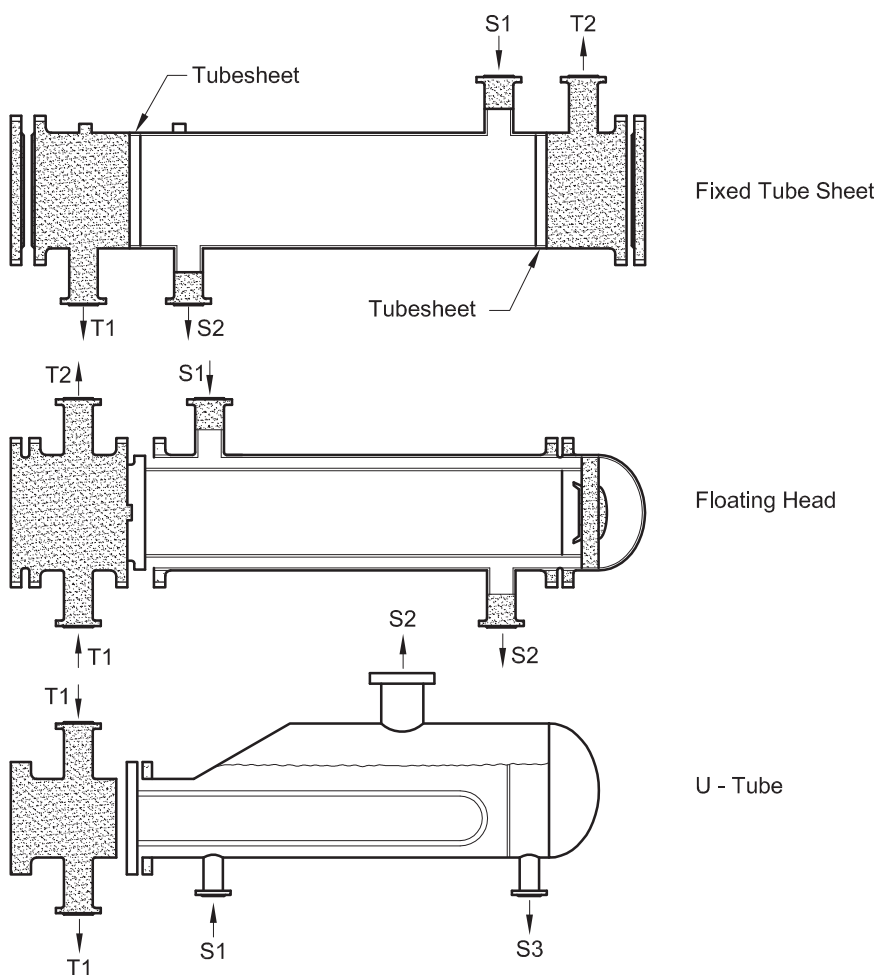


Figure 4.18 Basic types of shell-and-tube exchangers.



Figure 4.19 TEMA type AET.



Figure 4.20 TEMA type NEN.



Figure 4.21 TEMA type BEM.



Figure 4.22 TEMA type BES (vertical).



Figure 4.23 TEMA type BKM (double kettle).



Figure 4.24 TEMA type CKU.

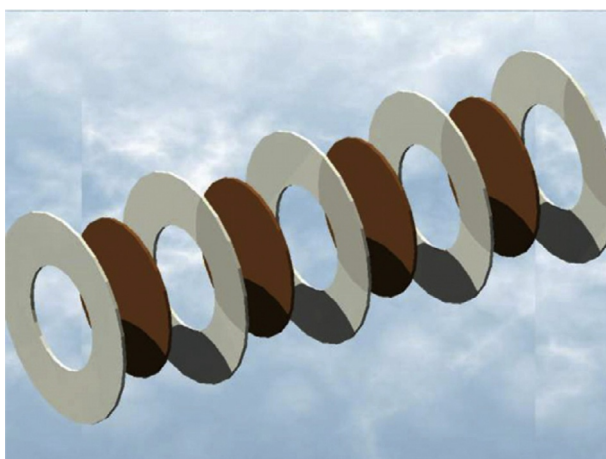


Figure 4.25 Shell-and-tube baffles—disk and donut.

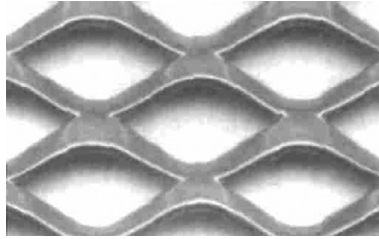


Figure 4.26 Shell-and-tube baffles—EM baffles.

4.2.19 U Tube (Hairpin)

4.2.19.1 Advantages

- Low cost—may be less expensive than fixed
- Can handle thermal expansion
- No internally gasketed joint is needed (good for high pressure)
- Can replace bundle

4.2.19.2 Disadvantages

- Tubes cannot be cleaned mechanically
- Individual tubes are not replaceable
- Fewer tubes can be installed in a given shell diameter

4.2.19.3 Heads

- B and U are most common

4.2.19.4 Uses

- High temperature
- Clean tube fluids (need to minimize contamination of tube fluids with higher pressure shell fluids)

4.2.20 Floating Head

4.2.20.1 Advantages

- Most versatile
- High differential temperature
- Can clean both tubes and heads mechanically
- Can replace individual tubes

4.2.20.2 Disadvantages

- Most expensive
- Higher cost (at least 25% over fixed)

- Internal gasket can leak
- Difficult to plug individual tubes

4.2.20.3 Heads

- T and W are least expensive (potential of shell leakage to atmosphere)
- S is medium cost
- P is most expensive

4.2.20.4 Uses

- Dirty, high temperature differential service

4.2.21 Placement of Fluid

Consider placing fluid through:

Tubes when:

- Special alloy materials are required for corrosion control and high temperatures
- Fluid is at high pressure (minimizes cost)
- Fluid contains vapors and noncondensable gases (better heat transfer)
- Fluid is scale forming (can ream out tubes)
- Fluids is toxic or lethal (minimizes leakage)

Shell when:

- Small pressure drops are desired (less Δp going through shell)
- Fluid is viscous (less Δp and better heat transfer)
- Fluid is nonfouling (harder to clean shell)
- Fluid requires condensing or boiling service (kettle design best)
- Fluid has low flow rate and is nonfouling (finned tubes can be used)

4.2.22 Mean Temperature Difference Correction Factor

The “ F ” factor is unity in the mean temperature difference (MTD) equation for double-pipe exchangers and counter-current flow shell-and-tube exchangers with an equal number of shell and tube passes. The way fluid flows through the exchanger will affect the MTD. The correction factor is a function of

- Number of tube passes
- Number of shell passes

It is rare that a configuration should be chosen where “ F ” is <0.8 .

[Figures 4.27](#) and [4.28](#) provide a means to estimate the “ F ” factor.

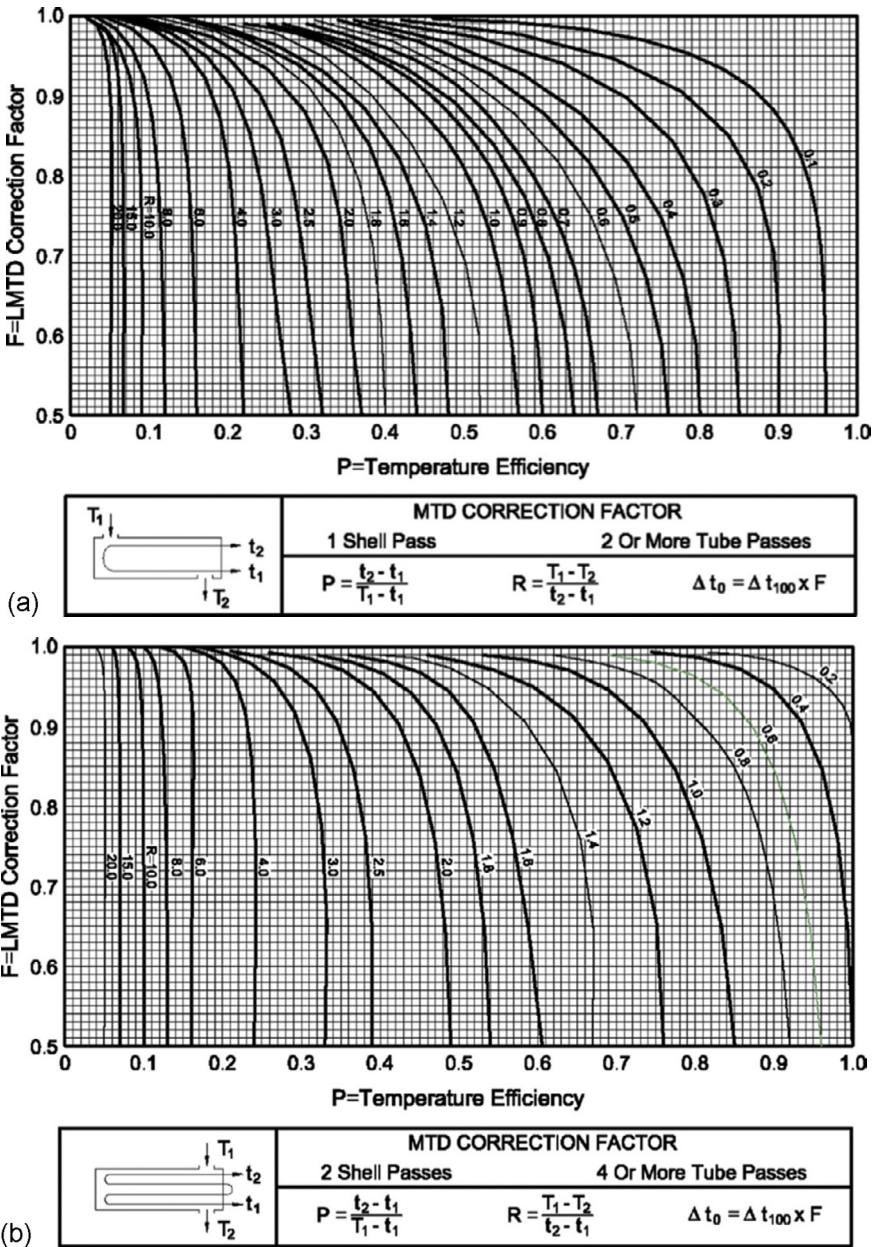


Figure 4.27 MTD geometry correction factors.

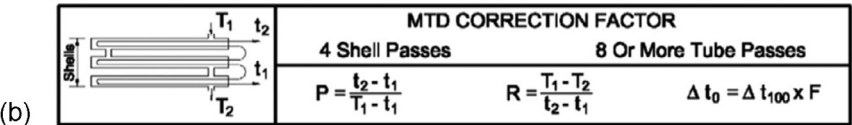
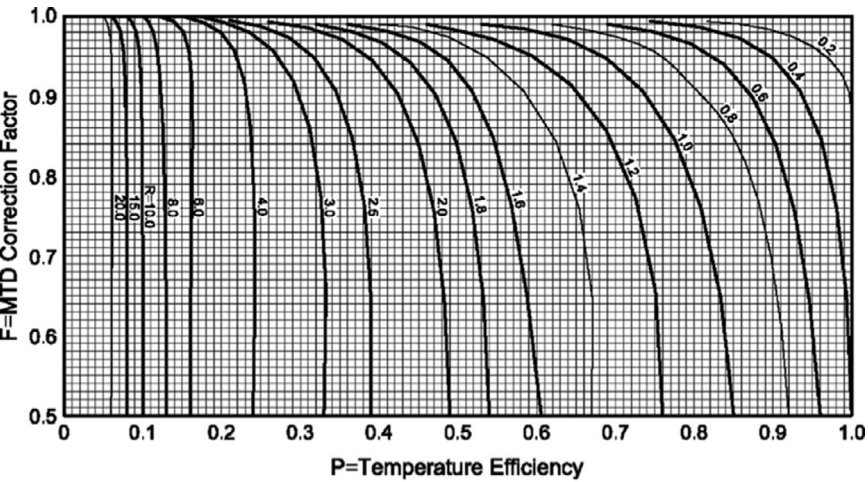
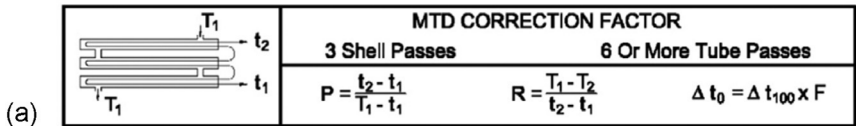
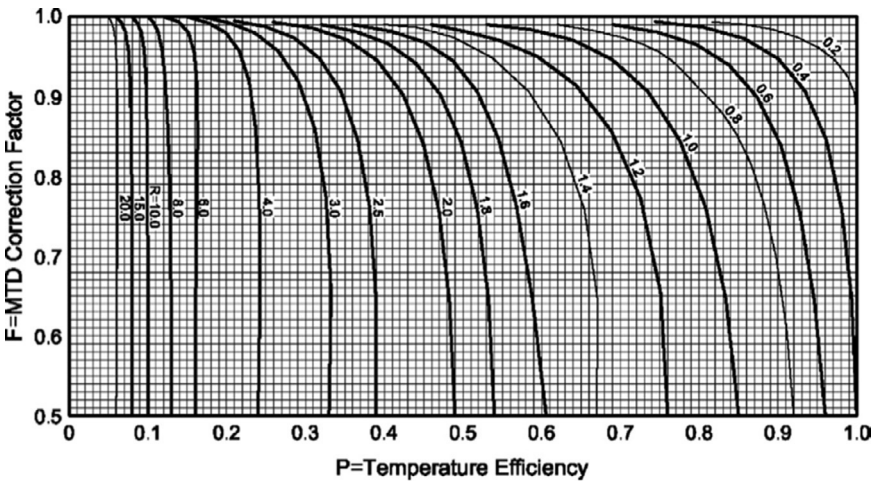


Figure 4.28 MTD geometry correction factors.

4.2.23 Corrected MTD

The mean temperature difference can be determined from the general equation:

$$\text{MTD} = F \left[\frac{\Delta T_1 - \Delta T_2}{l_N \left(\frac{\Delta T_1}{\Delta T_2} \right)} \right] \quad (4.1)$$

where T_1 = hot fluid inlet temperature, °F; T_2 = hot fluid outlet temperature, °F; T_3 = cold fluid inlet temperature, °F; T_4 = cold fluid outlet temperature, °F; ΔT_1 = larger temperature difference, °F; ΔT_2 = smaller temperature difference, °F; MTD = mean temperature difference, °F; F = correction factor (from Figures 4.27 and 4.28).

4.2.24 Heat Exchanger Specification Sheet

Figure 4.29 is an example of a TEMA heat exchanger specification sheet.

4.2.25 Sizing Procedures

The following steps indicate how to size a heat exchanger:

1. Calculate process heat duty
2. Determine which fluid will be in the shell and which fluid will be in the tube
3. Calculate/assume overall heat transfer coefficient
4. Select the
 - a. Number of shell passes
 - b. Number of tube passes
5. Calculate correction factor and corrected MTD
6. Choose tube diameter and tube length
7. Calculate number of tubes required from the following

$$N = \frac{q}{UA^1(\text{MTD})L} \quad (4.2)$$

where N = required number of tubes q . heat duty, Btu/h; U = overall heat transfer coefficient, Btu/h ft² °F; LMTD = corrected log mean temperature difference, °F; L = tube length, ft.; A^1 = tube external surface area per foot of length, ft²/ft (from Table 4.1).

1											Job No.	
2	Customer										Reference No.	
3	Address										Proposal No.	
4	Plant Location										Date	Rev.
5	Service of Unit										Item No.	
6	Size	Type	(Hor/Vert)		Connected		In	Parallel	Series			
7	Surf/Unit (Gross / Eff)	Sq	Ft	Shells/Unit	Surf/Shell (Gross/Eff.)				Sq	Ft		
8	PERFORMANCE OF ONE UNIT											
9	Fluid Allocation				Shell Side			Tube Side				
10	Fluid Name											
11	Fluid Quantity, Total				Lb/Hr							
12	Vapor (In/Out)											
13	Liquid											
14	Steam											
15	Water											
16	Noncondensable											
17	Temperature (In/Out)				°F							
18	Specific Gravity											
19	Viscosity, Liquid				Cp							
20	Molecular Weight, Vapor											
21	Molecular Weight, Noncondensable											
22	Specific Heat				Btu/Lb °F							
23	Thermal Conductivity				Btu Ft/Hr Sq Ft °F							
24	Latent Heat				Btu/Lb @ °F							
25	Inlet Pressure				Psig							
26	Velocity				Ft/S							
27	Pressure Drop, Allow./Calc.				Psi			/				
28	Fouling Resistance (Min.)											
29	Heat Exchanged				Btu/Hr: MTD (Corrected)			°F				
30	Transfer Rate, Service				Clean			Btu/Hr Sq Ft °F				
31	CONSTRUCTION OF ONE SHELL											
32					Shell Side			Tube Side			Sketch (Bundle/Nozzle Orientation)	
33	Design/Test Pressure				Psig			/				
34	Design Temperature				°F							
35	No. Passes per Shell											
36	Corrosion Allowance				In.							
37	Connections				In							
38	Size & Rating				Out							
39					Intermediate							
40	Tube No.	OD	In.	Thk (Min/Avg)	In.	Length	Ft	Pitch	In.	↔ 30 ↗ 60 ↘ 90 ↙ 45		
41	Tube Type				Material							
42	Shell	ID	OD		In.	Shell Cover		(Integ.) (Remov.)				
43	Channel or Bonnet				Channel Cover							
44	Tubesheet-Stationary				Tubesheet-Floating							
45	Floating Head Cover				Impingement Protection							
46	Baffles-Cross				Type	% Cut (Diarm/Area)		Spacing: c/c	Inlet	In.		
47	Baffles-Long				Seal Type							
48	Supports-Tube				U-Bend		Type					
49	Bypass Seal Arrangement				Tube-Tubesheet Joint							
50	Expansion Joint				Type							
51	ov' - Inlet Nozzle				Bundle Entrance		Bundle Exit					
52	Gaskets-Shell Side				Tube Side							
53	-Floating Head											
54	Code Requirements				TEMA Class							
55	Weight/Shell				Filled with Water			Bundle		Lb		
56	Remarks											
57												
58												
59												
60												
61												

Figure 4.29 TEMA shell-and-tube exchanger specification sheet.

Select a shell diameter that can accommodate the number of tubes required (Figure 4.30 and Tables 4.2–4.6). Figure 4.30 lists the total number of tubes, not the number of tubes per pass.

Multiple pass units have fewer total tubes due to partition plate. Floating head units have fewer tubes than fixed head types because heads and seals

Table 4.1 Characteristics of Tubing

Tube OD in.	B.W.G. Gauge	Thickness in.	Internal Area in.	FT External Surface per ft Length	FT Internal Surface per ft Length
1/4	22	0.028	0.0295	0.0655	0.0508
1/4	24	0.022	0.0333	0.0655	0.0539
1/4	26	0.018	0.0360	0.0655	0.0560
1/4	27	0.016	0.0373	0.0655	0.0570
3/8	18	0.049	0.0603	0.0982	0.0725
3/8	20	0.035	0.0731	0.0982	0.0798
3/8	22	0.028	0.0799	0.0982	0.0835
3/8	24	0.022	0.0860	0.0982	0.0867
1/2	16	0.065	0.1075	0.1309	0.0969
1/2	18	0.049	0.1269	0.1309	0.1052
1/2	20	0.035	0.1452	0.1309	0.1126
1/2	22	0.028	0.1548	0.1636	0.1162
5/8	12	0.109	0.1301	0.1636	0.1066
5/8	13	0.095	0.1486	0.1636	0.1139
5/8	14	0.083	0.1655	0.1636	0.1202
5/8	15	0.072	0.1817	0.1636	0.1259
5/8	16	0.065	0.1924	0.1636	0.1296
5/8	17	0.058	0.2035	0.1636	0.1333
5/8	18	0.049	0.2181	0.1636	0.1380
5/8	19	0.042	0.2298	0.1636	0.1416
5/8	20	0.035	0.2419	0.1963	0.1453
3/4	10	0.123	0.1825	0.1963	0.1262
3/4	11	0.120	0.2043	0.1963	0.1335
3/4	12	0.109	0.2223	0.1963	0.1393
3/4	13	0.095	0.2463	0.1963	0.1466
3/4	14	0.083	0.2679	0.1963	0.1529
3/4	15	0.072	0.2884	0.1963	0.1587
3/4	16	0.065	0.3019	0.1963	0.1623
3/4	17	0.058	0.3157	0.1963	0.1660
3/4	18	0.049	0.3339	0.1963	0.1707
3/4	20	0.035	0.3632	0.2618	0.1780
1	8	0.165	0.3526	0.2618	0.1754
1	10	0.134	0.4208	0.2618	0.1916
1	11	0.120	0.4536	0.2618	0.1990
1	12	0.109	0.4803	0.2618	0.2047
1	13	0.095	0.5153	0.2618	0.2121
1	14	0.083	0.5463	0.2618	0.2183
1	15	0.072	0.5755	0.2618	0.2241
1	16	0.065	0.5945	0.2618	0.2278
1	18	0.049	0.6390	0.2618	0.2361
1	20	0.035	0.6793	0.2618	0.2435

Continued

Table 4.1 Characteristics of Tubing—cont'd

Tube OD in.	B.W.G. Gauge	Thickness in.	Internal Area in.	FT External Surface per ft Length	FT Internal Surface per ft Length
11/4	7	0.180	0.6221	0.3272	0.2330
11/4	8	0.165	0.6648	0.3272	0.2409
11/4	10	0.134	0.7574	0.3272	0.2571
11/4	11	0.120	0.8012	0.3272	0.2644
11/4	12	0.109	0.8365	0.3272	0.2702
11/4	13	0.095	0.8825	0.3272	0.2775
11/4	14	0.083	0.9229	0.3272	0.2838
11/4	16	0.065	0.9852	0.3272	0.2932
11/4	18	0.049	1.042	0.3272	0.3016
11/4	20	0.035	1.094	0.3272	0.3089
11/2	10	0.134	1.192	0.3927	0.3225
11/2	12	0.109	1.291	0.3927	0.3356
11/2	14	0.083	1.398	0.3927	0.3492
11/2	16	0.065	1.474	0.3927	0.3587
2	11	0.120	2.433	0.5236	0.4608
2	12	0.109	2.494	0.5236	0.4665
2	13	0.095	2.573	0.5236	0.4739
2	14	0.083	2.642	0.5236	0.4801

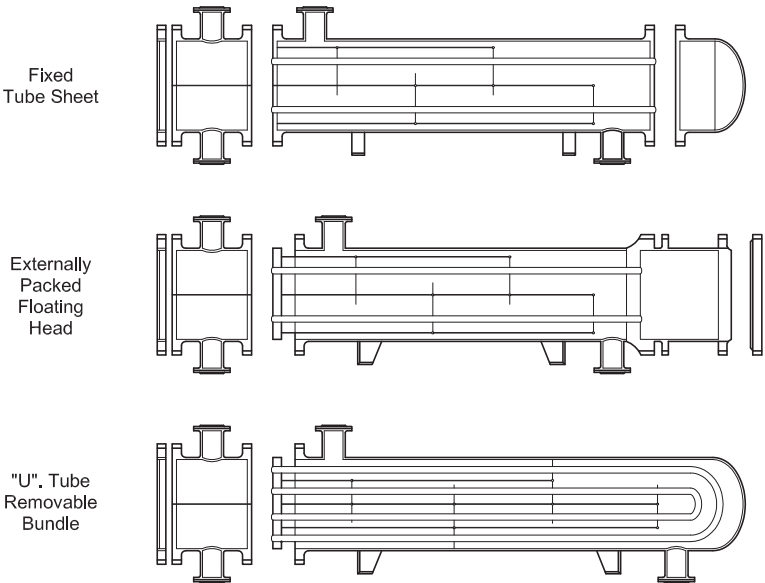


Figure 4.30 Heat exchanger tube count.

Table 4.2 Heat Exchanger Tube Count

Tube OD in.	B.W.G. Gauge	Thickness in.	Internal Area in.	FT External Surface per ft Length	FT Internal Surface per ft Length
1	14	0.083	0.5463	0.2618	0.2183
1	15	0.072	0.5755	0.2618	0.2241
1	16	0.065	0.5945	0.2618	0.2278
1	18	0.049	0.6390	0.2618	0.2361
1	20	0.035	0.6793	0.2618	0.2435
1 1/4	7	0.180	0.6221	0.3272	0.2330
1 1/4	8	0.165	0.6648	0.3272	0.2409
1 1/4	10	0.134	0.7574	0.3272	0.2571
1 1/4	11	0.120	0.8012	0.3272	0.2644
1 1/4	12	0.109	0.8365	0.3272	0.2702
1 1/4	13	0.095	0.8825	0.3272	0.2775
1 1/4	14	0.083	0.9229	0.3272	0.2838
1 1/4	16	0.065	0.9852	0.3272	0.2932
1 1/4	18	0.049	1.042	0.3272	0.3016
1 1/4	20	0.035	1.094	0.3272	0.3089
1 1/2	10	0.134	1.192	0.3927	0.3225
1 1/2	12	0.109	1.291	0.3927	0.3356
1 1/2	14	0.083	1.398	0.3927	0.3492
1 1/2	16	0.065	1.474	0.3927	0.3587
2	11	0.120	2.433	0.5236	0.4608
2	12	0.109	2.494	0.5236	0.4665
2	13	0.095	2.573	0.5236	0.4739
2	14	0.083	2.642	0.5236	0.4801

Table 4.3 Heat Exchanger Tube Count

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
3/4" OD Tubes on 1" □ Pitch								
5.047	12	12	12	12	6	4	3	2
6.065	21	16	16	16	16	12	4	4
7.981	37	34	32	32	28	24	12	10
10.02	61	60	52	52	52	52	22	20
12.00	97	88	88	81	78	76	34	34
13.25	112	112	112	97	94	88	45	44
15.25	156	148	148	140	132	124	64	60
17.25	208	196	188	188	178	172	88	84
19.25	250	249	244	241	224	216	112	108

Continued

Table 4.3 Heat Exchanger Tube Count—cont'd

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
21.25	316	307	296	296	280	276	138	134
23.25	378	370	370	356	344	332	170	166
25.00	442	432	428	414	406	392	200	194
27.00	518	509	496	482	476	468	236	230
29.00	602	596	580	570	562	548	277	272
31.00	686	676	676	658	640	640	320	312
33.00	782	768	768	742	732	732	362	360
35.00	896	868	868	846	831	820	418	406
37.00	1004	978	964	952	931	928	470	462
39.00	1102	1096	1076	1062	1045	1026	524	520
42.00	1283	1289	1270	1232	1222	1218	611	602
45.00	1484	1472	1456	1424	1415	1386	710	700
48.00	1701	1691	1670	1636	1634	1602	812	802
51.00	1928	1904	1888	1845	1832	1818	926	910
54.00	2154	2138	2106	2080	2066	2044	1042	1032
60.00	2683	2650	2636	2582	2566	2556	1298	1282
3/4" OD Tubes on 1" ♦ Pitch								
5.047	12	10	8	12	10	8	2	2
6.065	21	18	16	16	12	8	5	4
7.981	37	32	28	32	28	24	12	10
10.02	61	54	48	52	46	40	21	18
12.00	97	90	84	81	74	68	33	32
13.25	113	108	104	97	92	84	43	40
15.25	156	146	136	140	134	128	62	58
17.25	208	196	184	188	178	168	87	82
19.25	256	244	236	241	228	216	109	104
21.25	314	299	294	300	286	272	136	130
23.25	379	363	352	359	343	328	267	160
25.00	448	432	416	421	404	392	195	190
27.00	522	504	486	489	472	456	234	226
29.00	603	583	568	575	556	540	275	266
31.00	688	667	654	660	639	624	313	304
33.00	788	770	756	749	728	708	360	350
35.00	897	873	850	846	826	804	409	398
37.00	1009	983	958	952	928	908	464	452
39.00	1118	1092	1066	1068	1041	1016	518	508
42.00	1298	1269	1250	1238	1216	1196	610	596
45.00	1500	1470	1440	1424	1407	1378	706	692
48.00	1714	1681	1650	1644	1611	1580	804	788
51.00	1939	1903	1868	1864	1837	1804	917	902
54.00	2173	2135	2098	2098	2062	2026	1036	1018
60.00	2692	2651	2612	2600	2560	2520	1292	1272

Table 4.4 Heat Exchanger Tube Count
Shell ID (Inches) **Fixed Tube Sheet**

				Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
1" OD Tubes on 1-1/4" Δ Pitch								
5.047	8	6	4	7	4	4	0	0
6.065	14	14	8	10	10	4	2	2
7.981	26	26	16	22	18	16	7	4
10.02	42	40	36	38	36	28	13	12
12.00	64	61	56	56	52	48	22	18
13.25	75	76	72	73	72	60	28	26
15.25	110	106	100	100	98	88	43	38
17.25	147	138	128	130	126	116	57	52
19.25	184	175	168	170	162	148	76	68
21.25	227	220	212	212	201	188	96	88
23.25	280	265	252	258	250	232	116	110
25.00	316	313	294	296	294	276	135	128
27.00	371	370	358	355	346	328	161	152
29.00	434	424	408	416	408	392	189	182
31.00	503	489	468	475	466	446	222	212
33.00	576	558	534	544	529	510	254	246
35.00	643	634	604	619	604	582	289	280
37.00	738	709	468	696	679	660	330	316
39.00	804	787	772	768	753	730	370	356
42.00	946	928	898	908	891	860	436	418
45.00	1087	1069	1042	1041	1017	990	505	490
48.00	1240	1230	1198	1189	1182	1152	578	562
51.00	1397	1389	1354	1348	1337	1300	661	642
54.00	1592	1561	1530	1531	1503	1462	748	726
60.00	1969	1945	1904	1906	1979	1842	933	914
1" OD Tubes on 1-1/4" □ Pitch								
5.047	9	6	4	5	4	4	0	0
6.065	12	12	12	12	6	4	2	2
7.981	22	20	16	21	16	16	6	4
10.02	38	38	32	32	32	32	12	10
12.00	56	56	52	52	52	44	19	18
13.25	69	66	66	61	60	52	25	24
15.25	97	90	88	89	84	80	36	34
17.25	129	124	120	113	112	112	49	48
19.25	164	158	148	148	144	140	64	62
21.25	202	191	184	178	178	172	83	78

Continued

Table 4.4 Heat Exchanger Tube Count—cont'd

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
23.25	234	234	222	216	216	208	100	98
25.00	272	267	264	258	256	256	120	116
27.00	328	317	310	302	300	296	142	138
29.00	378	370	370	356	353	338	166	166
31.00	434	428	428	414	406	392	145	192
33.00	496	484	484	476	460	260	221	218
35.00	554	553	532	542	530	518	254	248
37.00	628	612	608	602	596	580	287	280
39.00	708	682	682	676	649	648	322	314
42.00	811	811	804	782	780	768	379	374
45.00	940	931	918	904	894	874	436	434
48.00	1076	1061	1040	1034	1027	1012	501	494
51.00	1218	1202	1192	1178	1155	1150	573	570
54.00	1370	1354	1350	1322	1307	1284	650	644
60.00	1701	1699	1684	1654	1640	1632	810	802

Table 4.5 Heat Exchanger Tube Count

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
1" OD Tubes On 1-1/4" Δ Pitch								
5.047	8	6	4	5	4	4	0	0
6.065	12	10	8	12	10	8	2	2
7.981	24	20	16	21	18	16	5	4
10.02	37	32	28	32	32	28	12	10
12.00	57	53	48	52	46	40	18	16
13.25	70	70	64	61	58	56	25	22
15.25	97	90	84	89	82	76	35	32
17.25	129	120	112	113	112	104	48	44
19.25	162	152	142	148	138	128	62	60
21.25	205	193	184	180	174	168	78	76
23.25	238	228	220	221	210	200	100	94
25.00	275	264	256	261	248	236	116	110
27.00	330	315	300	308	296	286	141	134

Table 4.5 Heat Exchanger Tube Count—cont'd

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
29.00	379	363	360	359	345	336	165	160
31.00	435	422	410	418	401	388	191	184
33.00	495	478	472	477	460	448	220	212
35.00	556	552	538	540	526	508	249	242
37.00	632	613	598	608	588	568	281	274
39.00	705	685	672	674	654	640	315	310
42.00	822	799	786	788	765	756	372	364
45.00	946	922	912	910	885	866	436	426
48.00	1079	1061	1052	1037	1018	1000	501	490
51.00	1220	1199	1176	1181	1160	1142	569	558
54.00	1389	1359	1330	1337	1307	1292	646	632
60.00	1714	1691	1664	1658	1626	1594	802	788

1-1/4" OD Tubes On 1-9/16" Δ Pitch

5.047	7	4	4	0	0	0	0	0
6.065	8	6	4	7	6	4	0	0
7.981	19	14	12	14	14	8	3	2
10.02	29	26	20	22	20	16	7	6
12.00	42	38	34	37	36	28	11	10
13.25	52	48	44	44	44	36	16	14
15.25	69	68	60	64	62	48	24	22
17.25	92	84	78	85	78	72	32	30
19.25	121	110	104	109	102	96	43	40
21.25	147	138	128	130	130	116	57	52
23.25	174	165	156	163	152	144	69	66
25.00	196	196	184	184	184	172	81	76
27.00	237	226	224	221	216	208	98	92
29.00	280	269	256	262	252	242	116	110
31.00	313	313	294	302	302	280	134	128
33.00	357	346	332	345	332	318	155	148
35.00	416	401	386	392	383	364	178	172
37.00	461	453	432	442	429	412	202	194
39.00	511	493	478	493	479	460	226	220
42.00	596	579	570	576	557	544	267	260
45.00	687	673	662	657	640	628	313	306
48.00	790	782	758	756	745	728	360	350
51.00	896	871	860	859	839	832	411	400
54.00	1008	994	968	964	959	940	465	454
60.00	1243	1243	1210	1199	1195	1170	580	570

Table 4.6 Heat Exchanger Tube Count

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
1-1/4" OD Tubes on 1-9/16" □ Pitch								
5.047	4	4	4	0	0	0	0	0
6.065	6	6	4	6	6	4	0	0
7.981	12	12	12	12	12	12	3	2
10.02	24	22	16	21	16	16	6	4
12.00	37	34	32	32	32	32	10	10
13.25	45	42	42	38	38	32	14	14
15.25	61	60	52	52	52	52	21	18
17.25	80	76	76	70	70	68	28	26
19.25	97	95	88	89	88	88	37	34
21.25	124	124	120	112	112	112	49	48
23.25	145	145	144	138	138	130	62	60
25.00	172	168	164	164	164	156	70	68
27.00	210	202	202	193	184	184	88	88
29.00	241	234	230	224	224	216	100	98
31.00	272	268	268	258	256	256	116	116
33.00	310	306	302	296	296	282	136	134
35.00	356	353	338	336	332	332	156	148
37.00	396	387	384	378	370	370	174	174
39.00	442	438	434	428	426	414	198	196
42.00	518	518	502	492	492	484	236	228
45.00	602	602	588	570	566	556	276	268
48.00	682	681	676	658	648	648	314	310
51.00	770	760	756	742	729	722	356	354
54.00	862	860	856	838	823	810	404	402
60.00	1084	1070	1054	1042	1034	1026	506	496
1-1/4" OD Tubes on 1-9/16" ◆ Pitch								
5.047	5	4	4	0	0	0	0	0
6.065	6	6	4	5	4	4	0	0
7.981	12	10	8	12	10	8	2	2
10.02	24	20	16	21	18	16	6	6
12.00	37	32	28	32	28	28	10	10
13.25	45	40	40	37	34	32	13	12
15.25	60	56	56	52	52	48	20	18
17.25	79	76	76	70	70	64	28	26
19.25	97	94	94	90	90	84	37	34
21.25	124	116	112	112	108	104	48	44

Table 4.6 Heat Exchanger Tube Count—cont'd

Shell ID (Inches)	Fixed Tube Sheet			Outside Packed Floating Head			U-Tube	
	No. of Passes			No. of Passes			No. of Passes	
	1	2	4	1	2	4	2	4
23.25	148	142	136	140	138	128	60	56
25.00	174	166	160	162	162	156	71	68
27.00	209	202	192	191	188	184	85	82
29.00	238	232	232	221	215	208	100	96
31.00	275	264	264	281	249	244	114	110
33.00	314	307	300	300	286	280	134	128
35.00	359	345	334	341	330	320	153	148
37.00	401	387	380	384	372	360	173	168
39.00	442	427	424	428	412	404	195	190
42.00	572	506	500	497	484	472	228	224
45.00	603	583	572	575	562	552	271	264
48.00	682	669	660	660	648	640	309	302
51.00	777	762	756	743	728	716	354	346
54.00	875	857	850	843	822	812	401	392
60.00	1088	1080	1058	1049	1029	1016	505	492

restrict the use of space. *U* tubes have the lowest number of tubes because of the space required for the tightest radius bend in the *U*-tube bundle. Once the number of tubes is determined, the flow velocity of fluid inside the tubes should be checked.

Example 4.1

Given

Geometry	4833/4-in. tubes, 20 ft long
<i>F</i>	1.0 (pure counterflow)
<i>W</i> _{hot}	100,000 lbs/h
<i>W</i> _{cold}	500,000 lbs/h
<i>C</i> _{p,hot}	0.77 Btu/lb-°F
<i>T</i> _{hot,in}	240 °F
<i>T</i> _{hot,out}	120 °F
<i>T</i> _{cold,in}	90 °F

Determine

Calculate the process heat duty, MTD and U

Solution**(1) Calculate area**

$$\begin{aligned} A &= n\pi dl \\ &= 483\pi(0.75/12)(20) \\ &= 1897\text{ft}^2 \end{aligned}$$

(2) Calculate process heat duty

$$\begin{aligned} Q_h &= WC_{p,\text{hot}}\Delta T_h \\ &= (100,000)(0.77)(240 - 120) \\ &= 9,240,000\text{Btu/h} \end{aligned}$$

(3) Calculate cold outlet temperature

$$Q_c = W_C C_{p,\text{cold}} \Delta T_C$$

Rearranging and solving for ΔT_c

$$\begin{aligned} \Delta T_C &= Q_C / (W_C C_{p,\text{cold}}) \\ T_{\text{cold,out}} &= T_{\text{cold,in}} + Q_C / (C_{p,\text{cold}}) \\ &= 90 + 9,240,000 / (50,000)(1.0) \\ &= 108.5^\circ\text{F} \end{aligned}$$

(4) Calculate LMTD

$$\begin{aligned} \text{LMTD} &= [(T_{\text{hot,in}} - T_{\text{cold,out}}) - (T_{\text{hot,out}} - T_{\text{cold,in}})] / \ln \\ &= [(T_{\text{hot,in}} - T_{\text{cold,out}}) / (T_{\text{hot,out}} - T_{\text{cold,in}})] \\ &= [(240 - 108.5) - (120 - 90)] / \ln[(240 - 108.5) / (120 - 90)] \\ &= 68.7^\circ\text{F} \end{aligned}$$

(5) Calculate U

$$Q = UAF(\text{LMTD})$$

Rearranging and solving for U

$$\begin{aligned} U &= Q / (AF(\text{LMTD})) \\ &= 9,240,000 / [(1897)(1.0)(68.7)] \\ &= 70.9\text{Btu/hft}^2\text{F} \end{aligned}$$



4.3 DOUBLE-PIPE EXCHANGERS

4.3.1 Overview

Double-pipe exchangers consist of a pipe or tube inside a pipe shell (Figure 4.31). They were developed to fit applications that are too small to use TEMA requirements. Tubes are often finned to yield additional surface area.

4.3.1.1 Advantages

- Inexpensive and readily available
- Can handle thermal expansion in U-bend construction

4.3.1.2 Disadvantages

- Limited to bare tube surface area of $<500 \text{ ft}^2$ (maximum 1000 ft^2)
- Cannot remove tube
- Cannot clean tube and shell mechanically

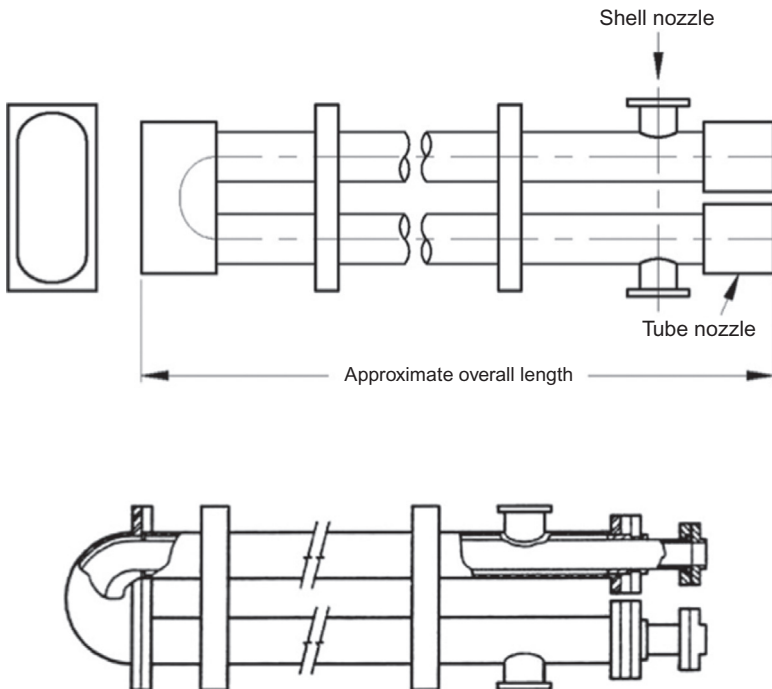


Figure 4.31 Double-pipe exchanger.

4.3.1.3 Uses

- Clean fluids
- Low heat transfer area required
- High temperature high pressure (500 psig)

4.3.2 Two Shells Joined at One End Through a "Return Bonnet" (Figure 4.32)

Shell-side fluid flows in series in each of two shells, which results in a more compact exchanger.

4.3.3 Hairpin Exchanger (Variation of U Tube)

4.3.3.1 Double Pipe

- Bare tube
- Longitudinal high fin

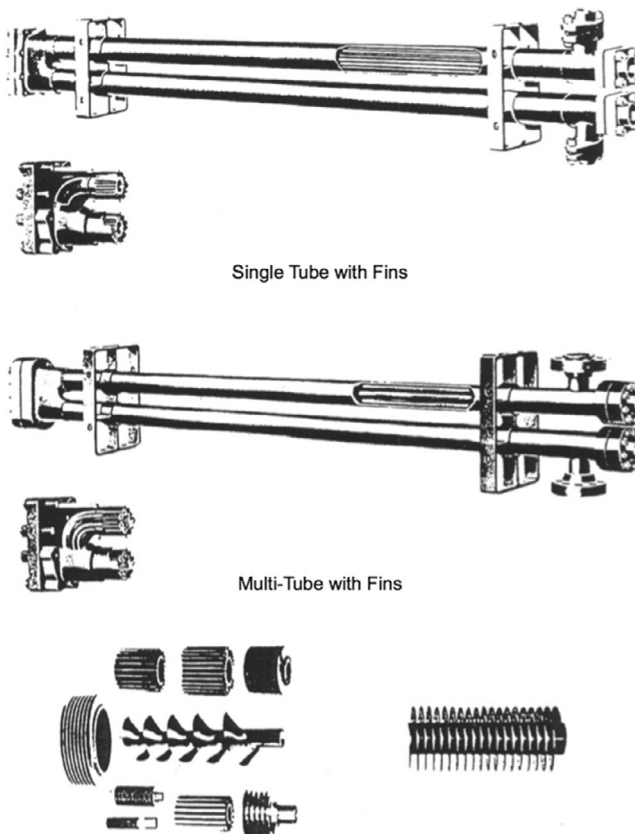


Figure 4.32 Details of finned tubes and double-pipe exchangers.

4.3.3.2 Multitube

Multiple small tubes bent into “U” shapes in place of a single tube allow for more surface area.

- Bare tube
- Longitudinal high fin or low fin

4.3.4 Design of Finned Units

The design of finned units is similar to other exchangers. The maximum velocity is limited by erosion, vibration, and pressure drop. See also [Figures 4.33–4.37](#).

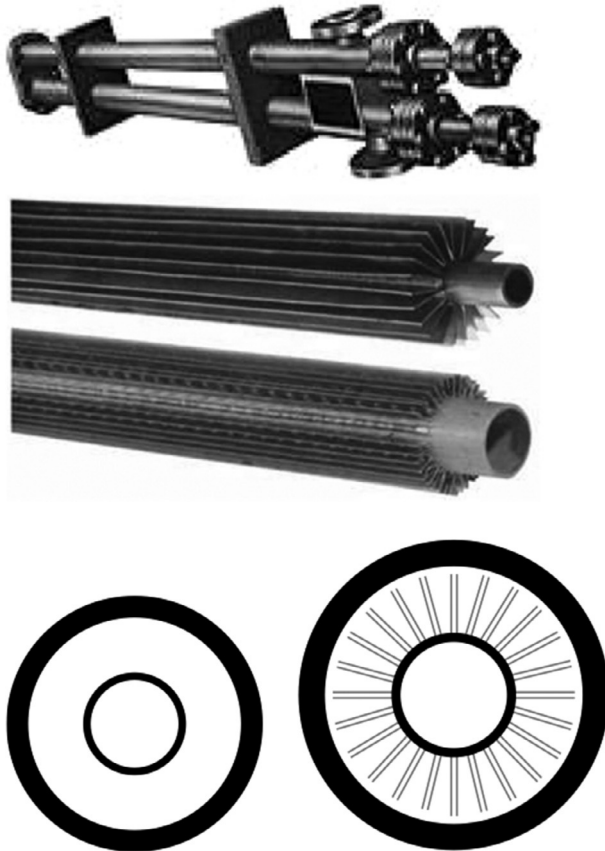
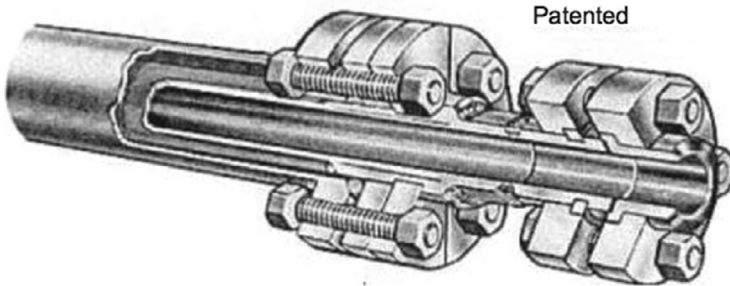
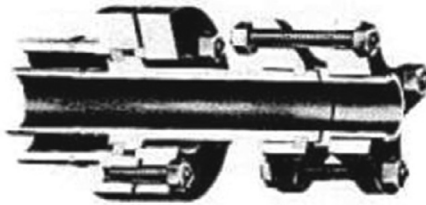


Figure 4.33 Hairpin double-pipe exchanger; top—bare tube; bottom—longitudinal high fin.

External Split Flange Design



Internal Split Ring Design



Taper-Lok[®] Tubeside Closure

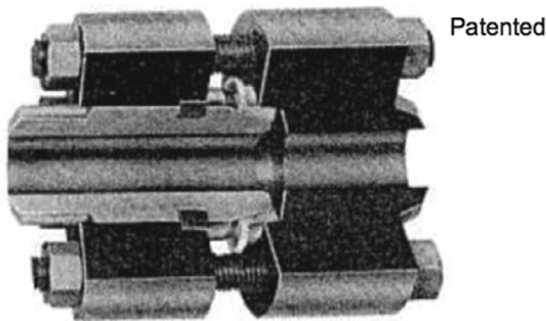


Figure 4.34 Hairpin double-pipe closures.



4.4 PLATE-FIN EXCHANGERS

4.4.1 Overview

Plate-fin heat exchangers are stacked, and the assemblies are placed into a vacuum-brazing furnace wherein the assembly is converted into an integral, solid structure (see [Figure 4.38](#)). During the process, an Al-Mg-Si alloy

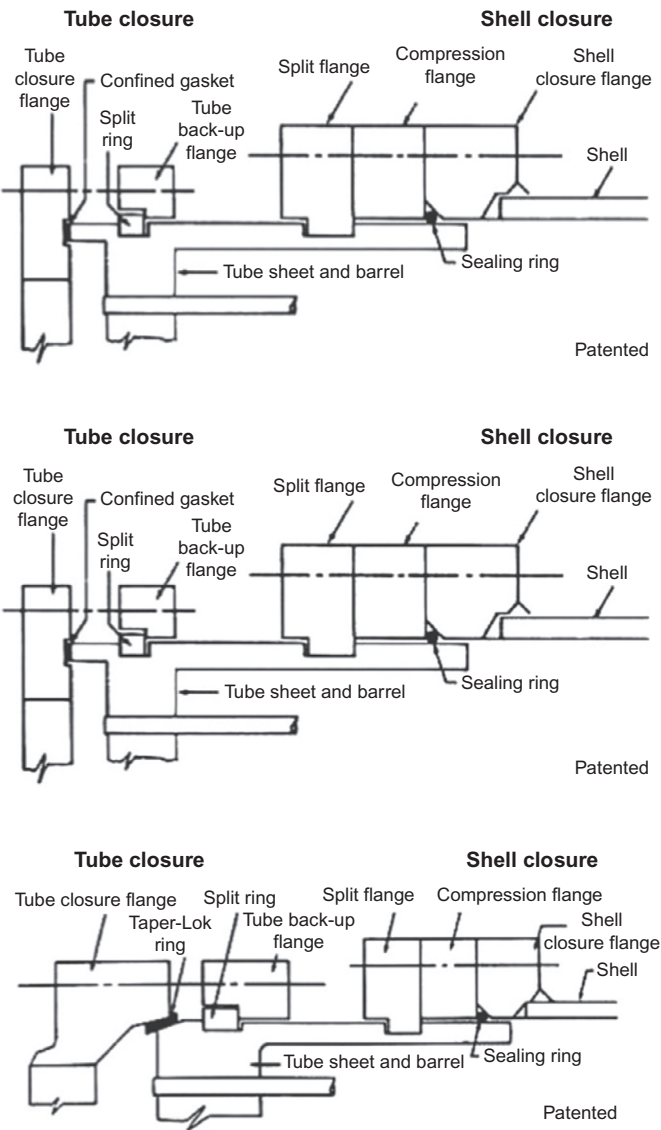


Figure 4.35 Hairpin multitube closures.

coating on the parting sheets melts and bonds the Al alloy fins, parting sheets, and bars together to form a core block. The melting temperature of the brazing alloy is within 50 °F (28 °C) of the melting temperature of the base Al alloys. The process is a very precise and highly protected process.

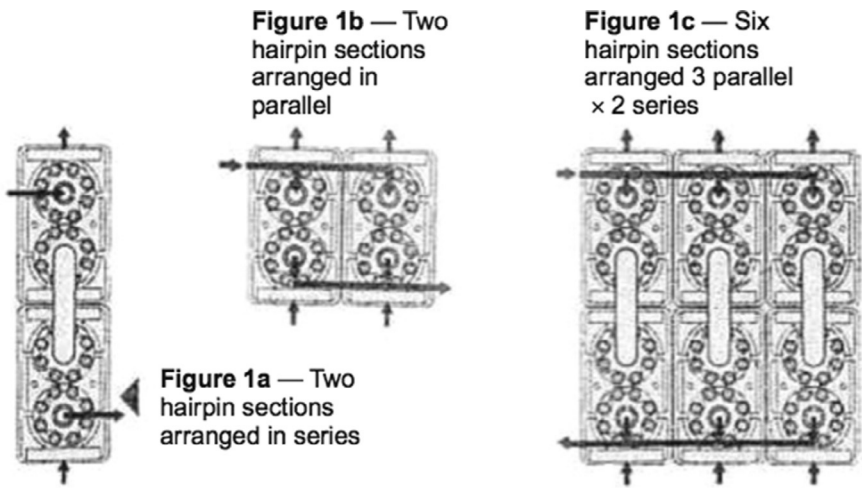


Figure 1 — Typical arrangement of sections to meet specific duties

Flexibility For Future Needs

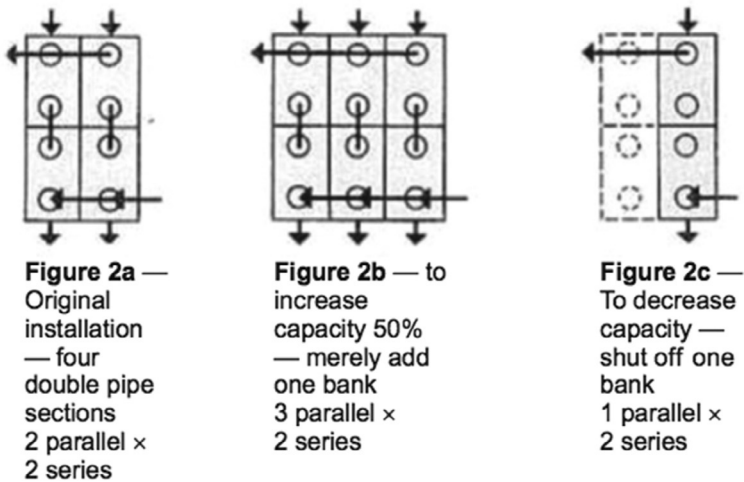


Figure 2 — How hairpin sections are changed to meet new requirements

Figure 4.36 Double-pipe hairpin banks.

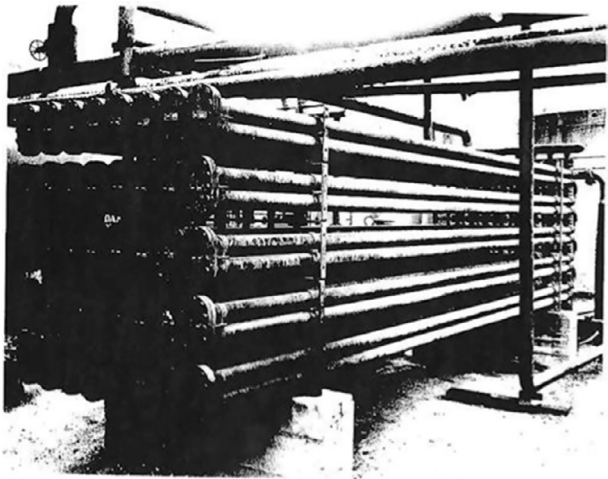


Figure 4.37 Stacked double-pipe exchanger.

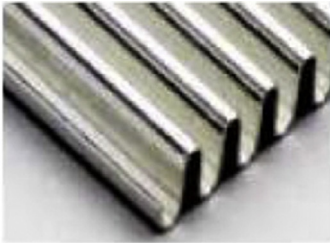


Figure 4.38 Vacuum-brazing process.

Braised aluminum plate fins are used in low-temperature (cryogenic) gas-processing services.

These plate fins can handle up to 10 fluids in a single exchanger, which is called a “cold box.” The “cold box” is composed of alternating layers of corrugated fins and flat separator sheets called parting sheets. A stack of fins and parting sheets make up this heat exchanger. Fins are constructed in four pattern types (Figure 4.39).

- Plain
- Perforated
- Herringbone
- Serrated



PLAIN • A sheet of metal with corrugated fins at right angles to the plates.



PERFORATED • A plain fin constructed from perforated material.



HERRINGBONE • Made by displacing the fins sideways at regular intervals to produce a zig-zag effect.



SERRATED • Made by simultaneously folding and cutting alternative sections of fins. These fins are also known as the lanced or multi-entry pattern.

Figure 4.39 Fin patterns for a brazed aluminum plate-fin exchanger.

4.4.2 Dependent on Application Requirements

The number of layers, types of fins, stacking arrangement, and the stream circuiting will vary depending on application requirements. Fluid configurations include cross-flow, counter-flow, and crosscounter-flow. See also [Figures 4.40–4.54](#).



4.5 PLATE-AND-FRAME EXCHANGERS

4.5.1 Overview

Plate-and-frame exchangers consists of layers of thin ribbed alloy plates, spaced about $\frac{1}{4}$ -in. apart, compressed together between gaskets (see [Figure 4.55](#)). Fluid flows between alternate sets of plates, and exchanges heat with the fluid flowing in the opposite direction. Distribution of hot and cold fluids to alternate plate flow channels is achieved by the gasketing pattern around each port. Alignment is achieved by top and bottom carrying bars and by slots in each plate. Fouling tendency between plates is the same as in tubes and through shell side at equal shell side at equal shear (for the same pressure drop per unit of heat transfer surface). Titanium plates are commonly used in cooling water/seawater/closed-loop water service on offshore platforms.



Figure 4.40 Brazed aluminum plate-fin exchangers.



Figure 4.41 Brazed aluminum plate-fin exchangers.

4.5.2 Advantages

- Compact in size and can obtain close temperature approaches
- Lighter and smaller than comparable shell-and-tube heat exchangers
- Allow full counter-current flow and thus no MTD correction is required
- Plates are easier to repair than fins

4.5.3 Disadvantages

- Less efficient, limited to
 - Moderate temperature [$<250\text{ }^{\circ}\text{F}$ ($<121\text{ }^{\circ}\text{C}$)]
 - Moderate pressure [$<200\text{ psig}$ ($<13.79\text{ barg}$)] due to gasketed service
- Should be used only where excessive fouling is not expected
- Requires superior metallurgy

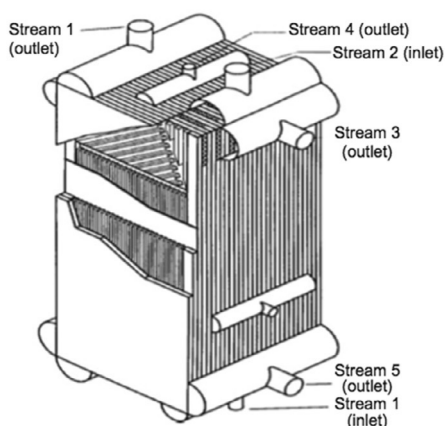
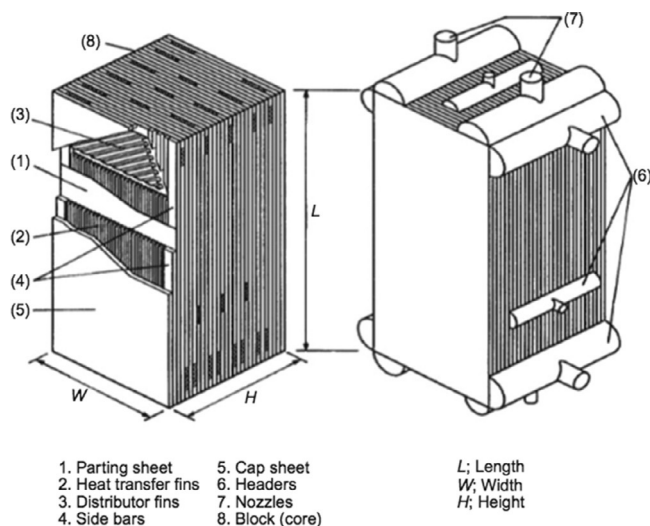


Figure 4.42 Braze aluminum plate-fin construction.

Gaskets can deteriorate in hydrocarbon service, offer poor fire resistance, and are expensive.

Arrangement of gasketed, pressed metal plates is aligned on carrying bars and is secured between two covers by compression bolts (Figure 4.56). Gaskets provide increased surface area.

Plates are made of thin, pressed metal and are resistant to corrosive attack. stainless steel (SS), monel, titanium, aluminum, and bronze are ideal for seawater/brackish services. Gasket material must be compatible with service.

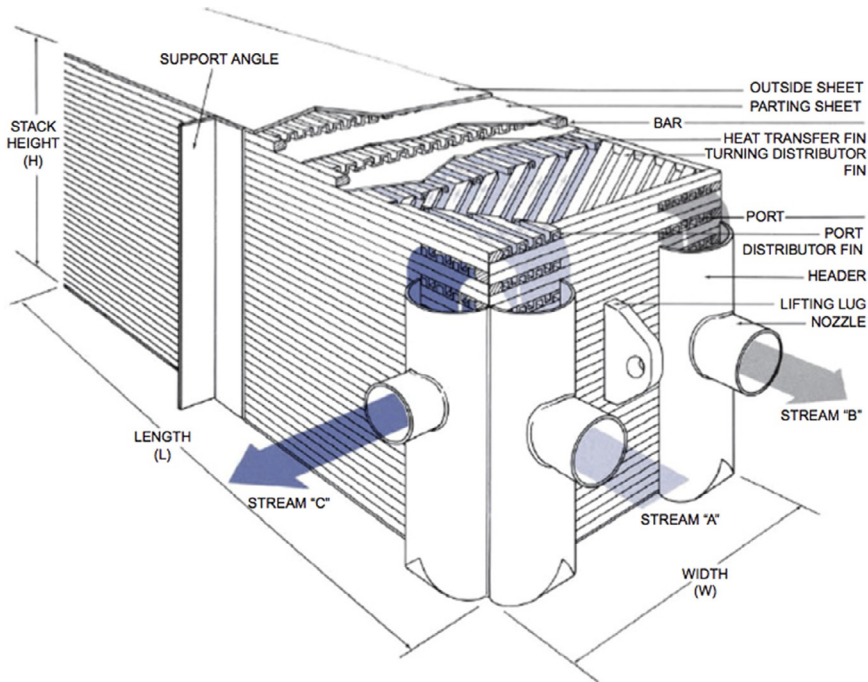


Figure 4.43 Brazed aluminum plate-fin construction.

4.5.4 Plates

Plates may be high-theta or low-theta plates. High-theta plates offer better heat transfer and higher pressure drop. Multiple plate styles may be present in a single unit. Different combinations of plates yield different results when used together. See also [Figures 4.57–4.68](#).

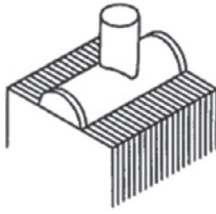


4.6 INDIRECT-FIRED HEATERS

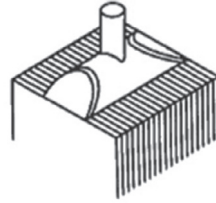
In an indirect-fired heater, hot combustion gas and flame heats an intermediate liquid, which, in turn, heats a fluid flowing through a coil or a series of tubes.

4.6.1 Advantages

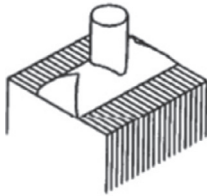
- Maintains a constant temperature over a long period of time
- Has proven safe, reliable, and convenient to use



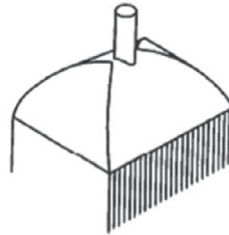
Standard header with flat ends



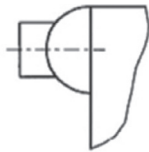
Header with inclined ends



Header with mitred ends



Dome header with mitred ends



Radial nozzle



Inclined nozzle



Tangential nozzle

Figure 4.44 Brazed aluminum plate-fin exchanger construction—headers.

4.6.2 Disadvantage

- Requires several hours to reach the desired temperature after it has been out of service

4.6.3 Intermediate Liquid

The intermediate liquid must be stable at atmospheric pressure and at maximum temperature. Depending on temperature level required, the fluid may use a water heating fluid medium. The transfer of heat between the fire tube and the fluid being heated is by natural convection. This limits the rate of heat flux per unit area and is rarely used to produce outlet temperatures

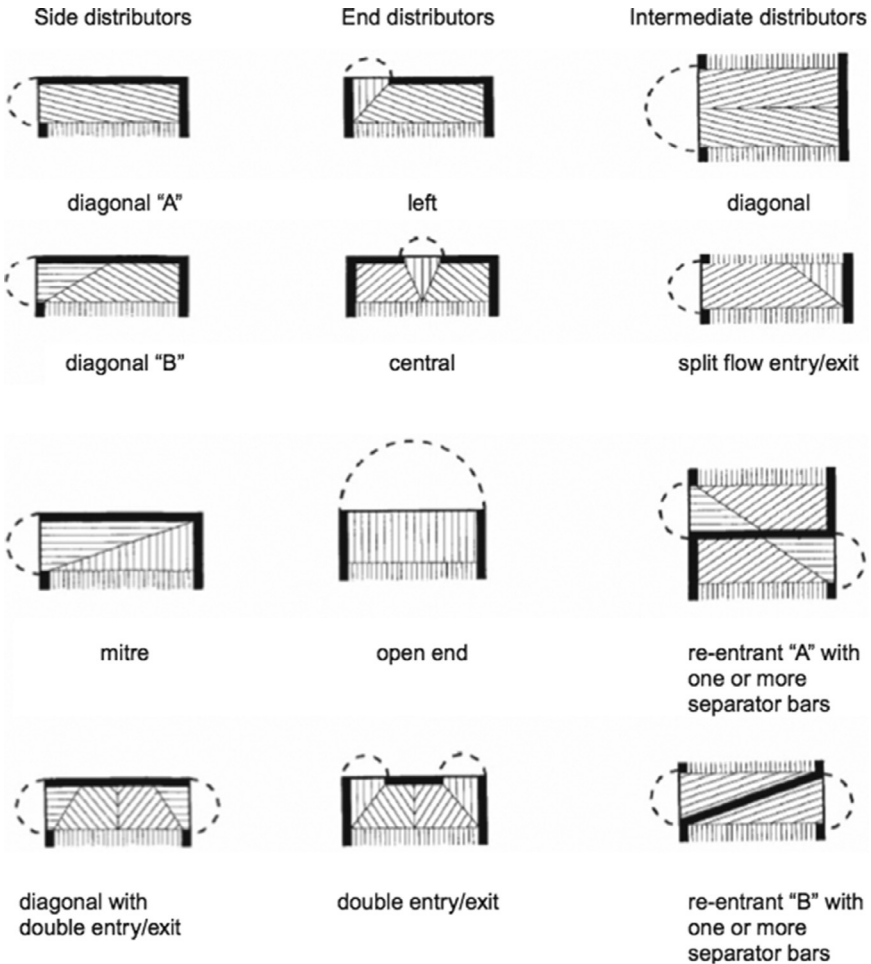


Figure 4.45 Brazed aluminum plate-fin construction—distributors.

above 500 °F. Intermediate fluids are primarily used to heat oil and gas in production operations where heat loads are not large. For example, line heaters in natural gas service (Figure 4.69), or reboilers in amine service (Figure 4.70).

In glycol and oil stabilization units, where the intermediate liquid is being boiled, there is no fluid coil or tube. These units usually consist of (Figure 4.71)

- Fire tube (usually fired by gas)
- Coil for fluid to be heated (Figure 4.72)
- A heat transfer fluid for immersion

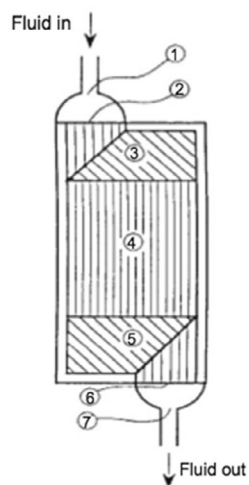
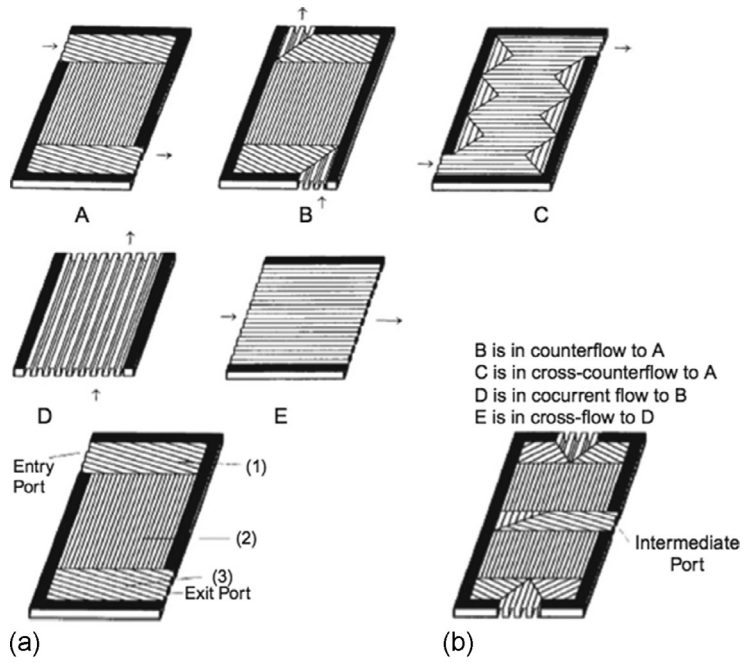


Figure 4.46 Brazed aluminum plate-fin construction—layout.

The designer must specify

- Heat duty
- Size of fire tube
- Coil diameter, length, and wall thickness

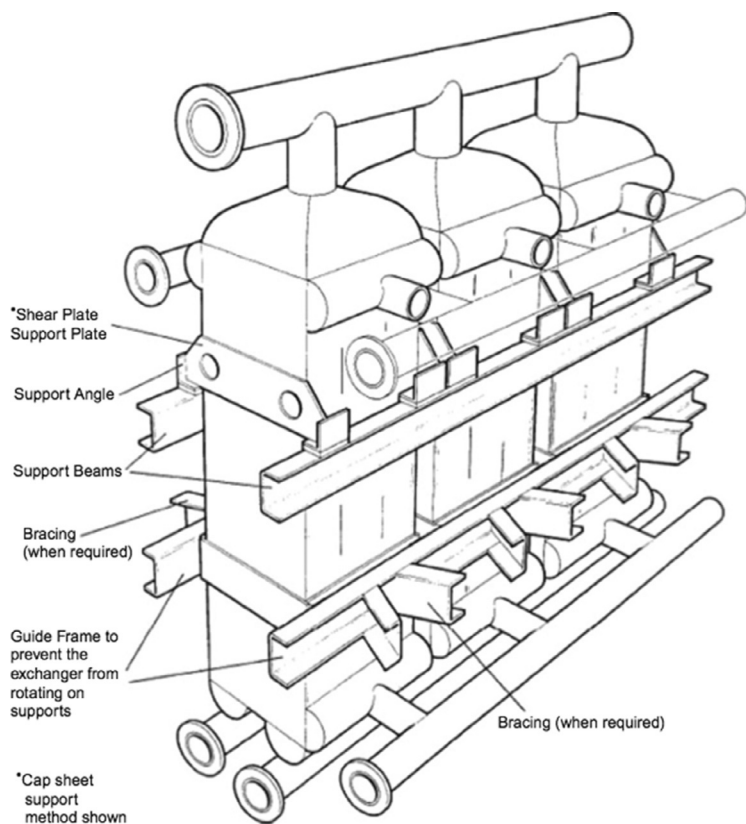


Figure 4.47 Brazed aluminum plate-fin assemblies.

4.6.4 Sizing Considerations

To determine heat duty required, we must know the following:

- Maximum gas, water, and oil/condensate expected in the heater
- Pressure and temperatures of heater inlet and outlet

The outlet temperature on gas “line” heaters depends on the hydrate formation temperature of the required heat duty. Special operating conditions must be considered in sizing. For example, start-up of a shut-in well may require additional heater capacity over steady-state requirements.

4.6.5 Heat Duty

The heat duty should be checked for various combinations of inlet temperature, pressure, flow rate, and outlet temperature and pressure to determine



Figure 4.48 Brazed aluminum plate-fin assemblies.

the most critical combination. This is best calculated using a process simulation program. The program accounts for phase changes as fluid passes through the choke (line heaters). It also balances enthalpies and change in temperature across the choke (line heaters). One must remember that the flow through a choke is instantaneous, no heat is absorbed or lost, but there is a temperature change. Flow through the coils is a constant pressure process. The change in enthalpy is equal to the heat absorbed, and the heat absorbed changes the temperature of the gas.



Figure 4.49 Brazed aluminum plate-fin pressure vessels.

The heat duty can be calculated from the general heat duty equation for multiphase streams

$$q = q_g + q_0 + q_w + q_l \quad (4.3)$$

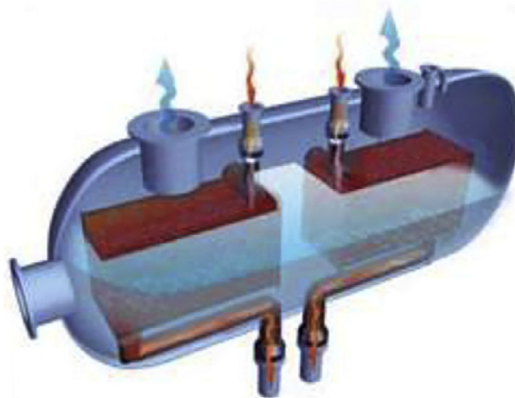
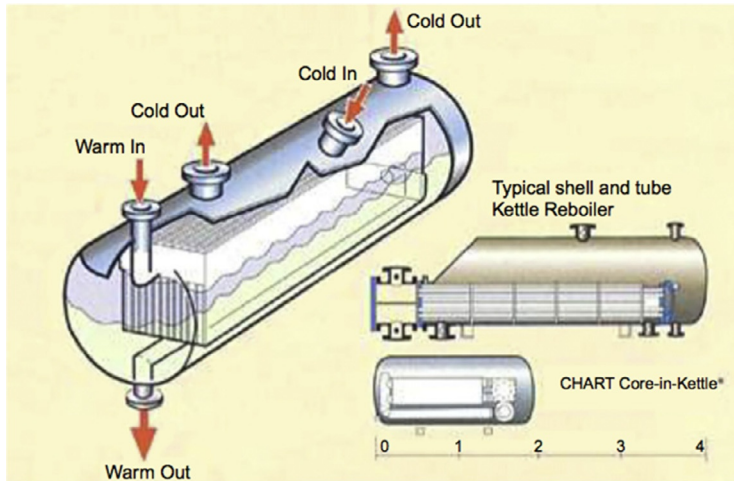


Figure 4.50 Brazed aluminum plate-fin core-in kettles.

where $q_g = 41.7 (\Delta t) C_g Q_g$; $q_0 = 14.6(SG)(\Delta t) C_g C_g$; $q_w = 14.6 (\Delta t) Q_w$; $q_{lost} = UA (190 - t_{amb})$, $= 0.10(q_g + q_0 + q_w)$; $\Delta t = T_{out} - T_{in} + \Delta T$ (due to pressure drop).

Calculate from enthalpy changes

$$q = h_{out} - h_{in} + q_{lost} \quad (4.4)$$

where h_{out} = enthalpy at P_{out} , T_{out} ; h_{in} = enthalpy at P_{in} , T_{in} .

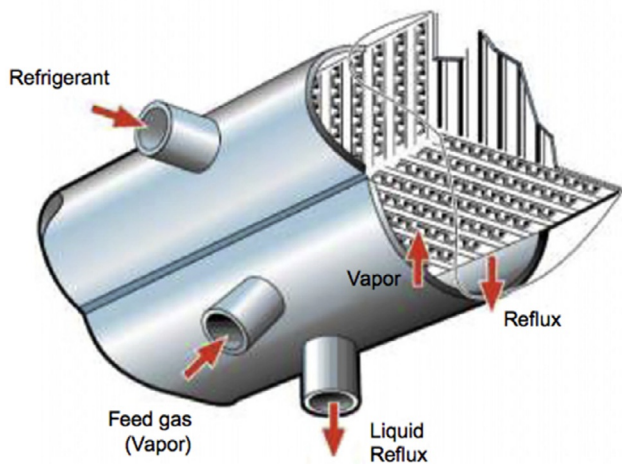
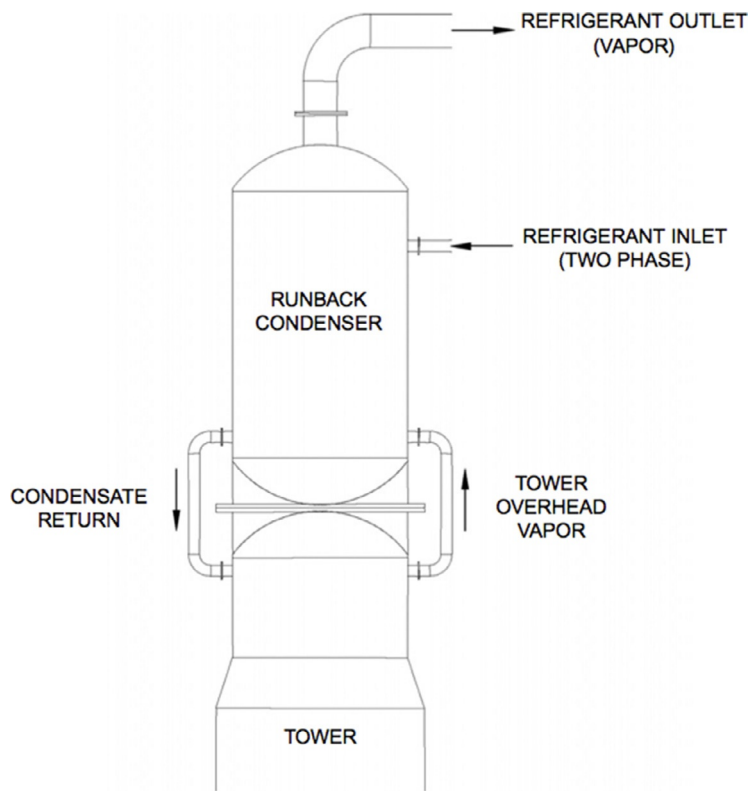
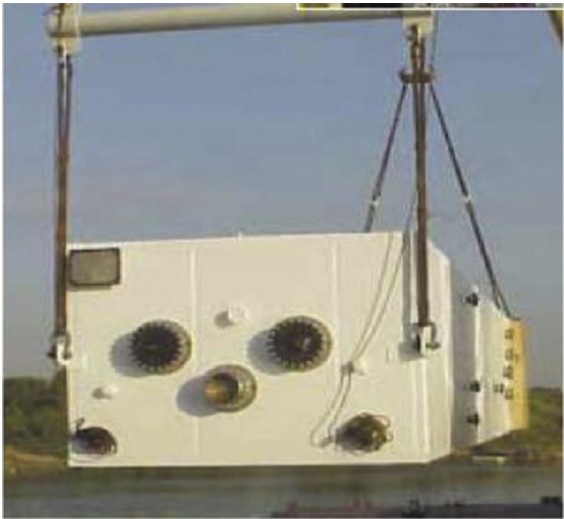
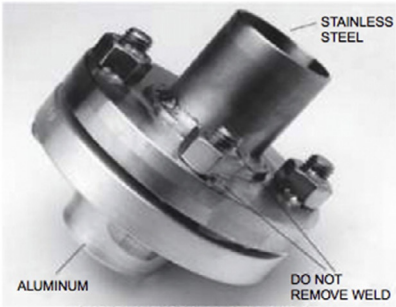


Figure 4.51 Brazed aluminum plate-fin reflux condensers.



STANDARD TRANSITION JOINT



FLANGE REINFORCED TRANSITION JOINT

Figure 4.52 Brazed aluminum plate-fin in cold boxes.

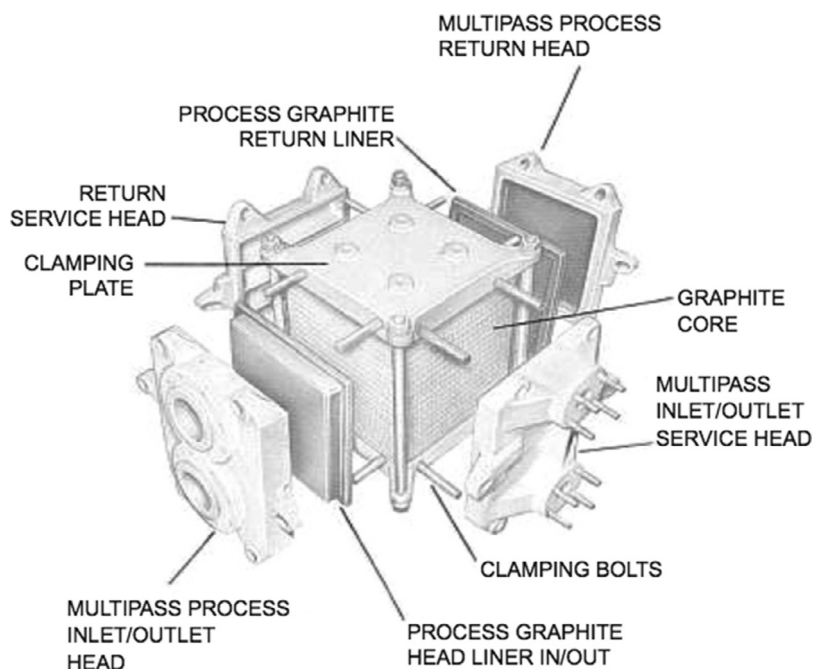


Figure 4.53 Diffusion bonded blocks.

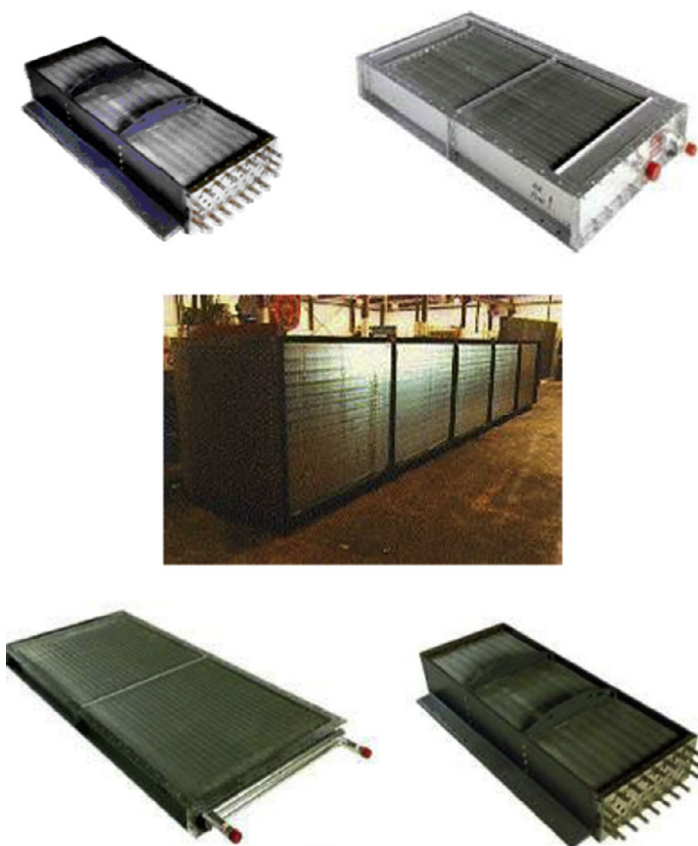


Figure 4.54 Plate-fin tube coil configuration.

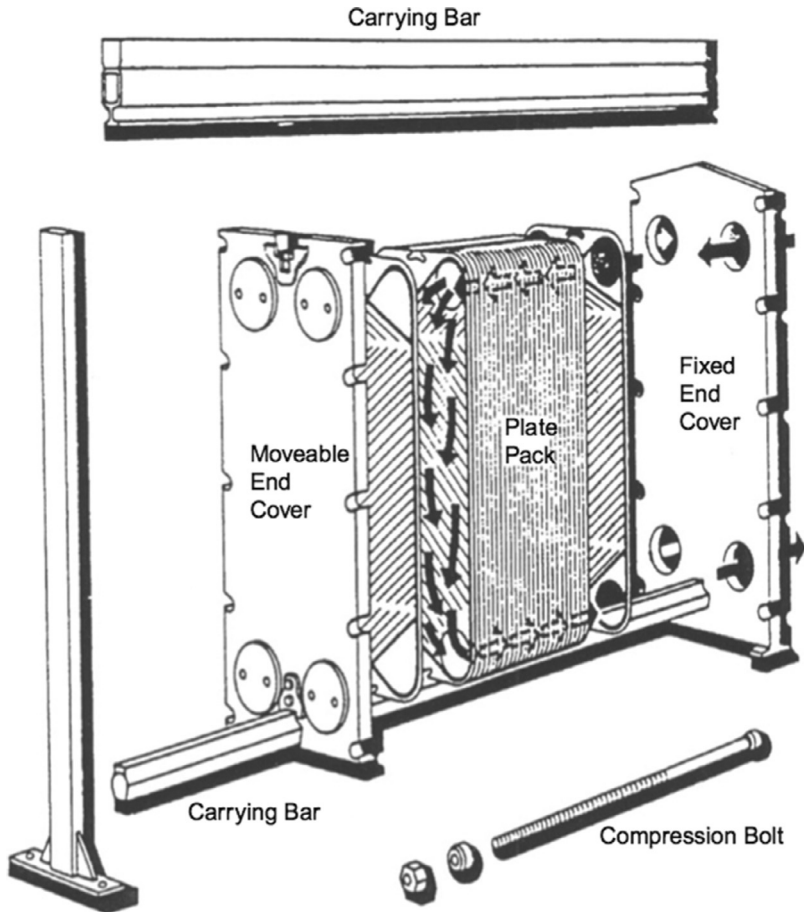


Figure 4.55 Plate-and-frame exchanger.

4.6.6 Sizing Fire Tubes

The required area of a fire tube with a flux rate of $10,000 \text{ Btu/h ft}^2$, recommended heat flux for maximum life, can be calculated from the following equation:

$$L = 3.8 \times 10^{-4} \left(\frac{q}{d} \right) \quad (4.5)$$

where L = fire tube length, ft; q = total heat duty, Btu/h; d = fire tube diameter, in.

The burner should be selected from a standard “off-the-shelf” unit. Manufacturers normally have standard diameters and lengths for different

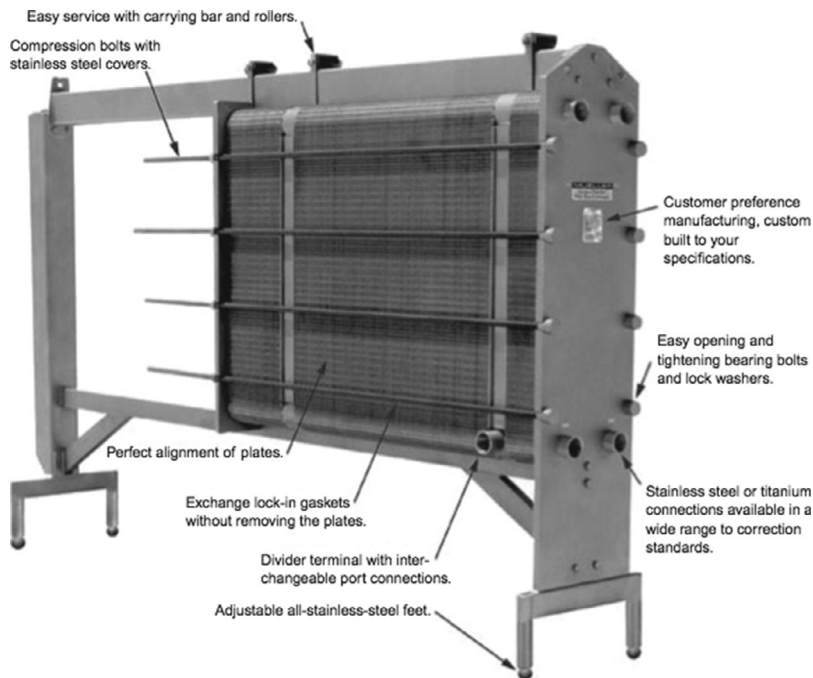


Figure 4.56 Gasketed plate-and-frame exchanger.

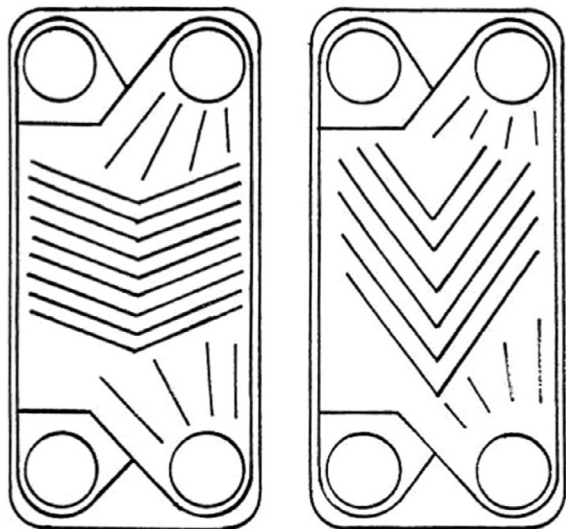


Figure 4.57 Traditional gasketed plate-and-frame design.

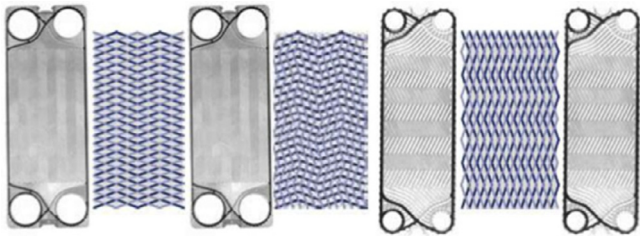


Figure 4.58 Multiple plate styles in a traditional plate-and-frame exchanger.

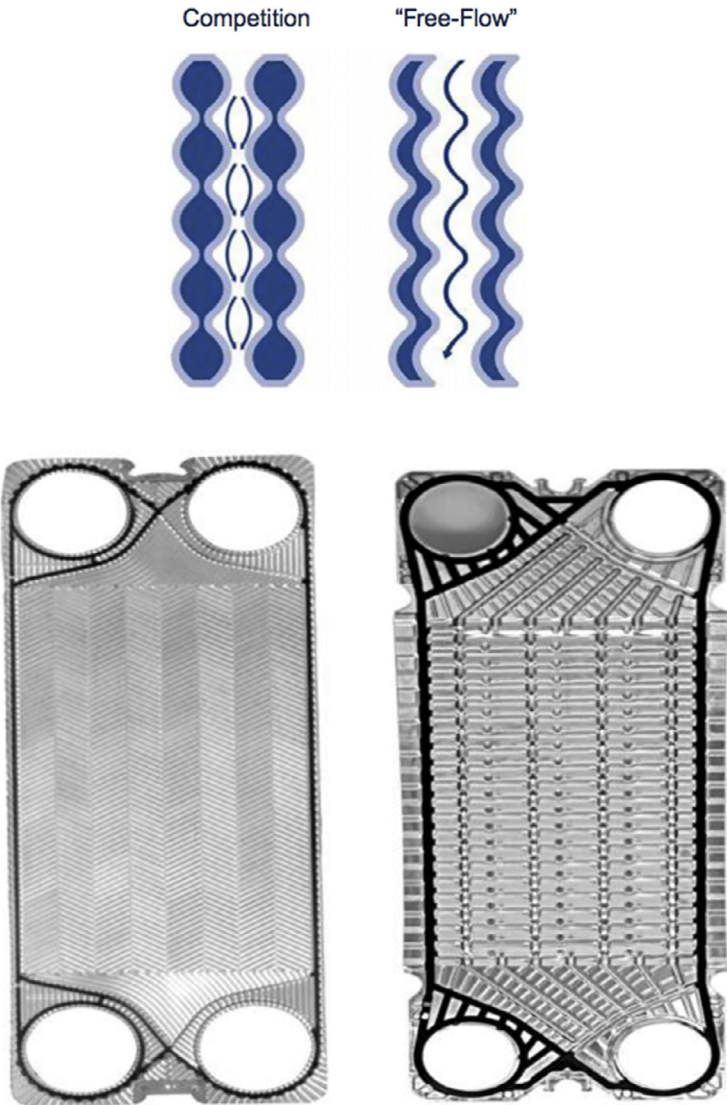


Figure 4.59 Traditional gasketed plate-and-frame with wide gap plate design.

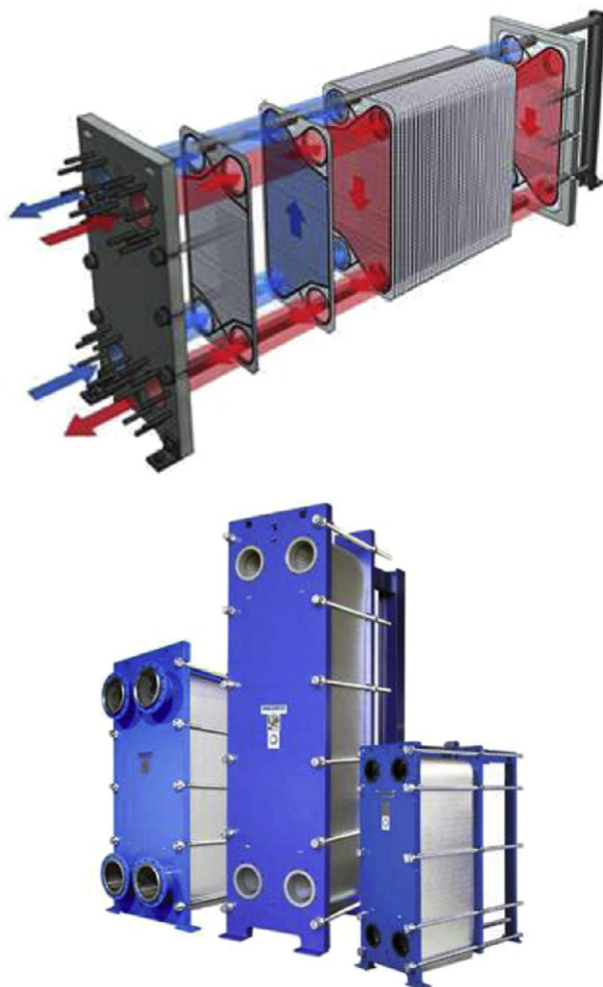


Figure 4.60 Traditional gasketed plate-and-frame design.

size fire tube ratings (Table 4.7). In order to choose the coil length and diameter, a temperature must first be chosen upstream of the choke.

4.6.7 Coil Sizing

4.6.7.1 Calculate the Coil Temperature

The first step in sizing the coil is to choose a temperature (upstream of choke). The greater the temperature difference between the gas and the bath, the smaller the coil area required. The bath temperature is constant at 190 °F (88 °C). The gas will be coldest downstream of choke.

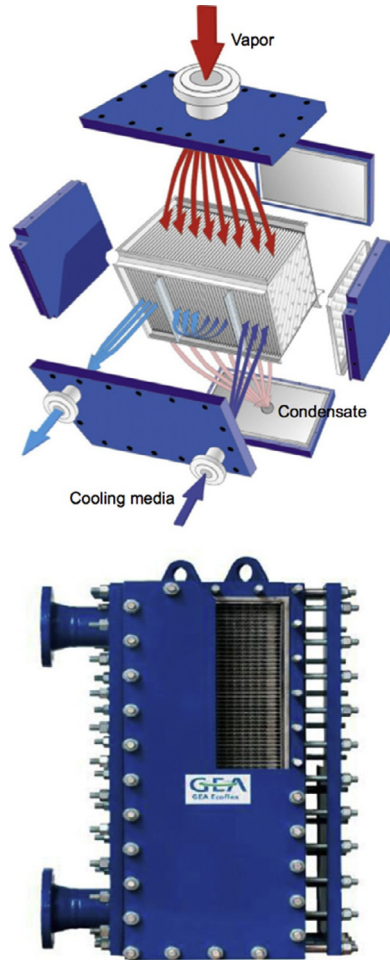


Figure 4.61 Welded “printed circuit” plate-and-frame design.

Therefore, the shortest total coil length ($L_1 + L_2$) will occur when L_1 is small. If L_1 becomes too long, consider using a glycol/water mixture or other heating medium, or raise the bath temperature. Ideally, try to keep T_2 above 50°F (10°C) to minimize plugging and above -20°F (-29°C) to avoid more costly steel. When determining the coil diameter, consider the pressure drop and maximum velocity that could occur. Typically, the pressure drop will not govern. The maximum velocity should be the lowest of the following:



Figure 4.62 Welded plate-and-frame.

- 60 fps for noise
- 50 fps in CO₂ service where a corrosion inhibitor is used
- Erosional velocity

The erosional velocity will always govern. The erosional velocity can be calculated from the following equation:

$$V_e = \frac{c}{\rho_m^{1/2}} \quad (4.6)$$

The wall thickness can be determined from ASME B31.8, B31.3, ASME Pressure Vessel Code, or API 12K. Before choosing a wall

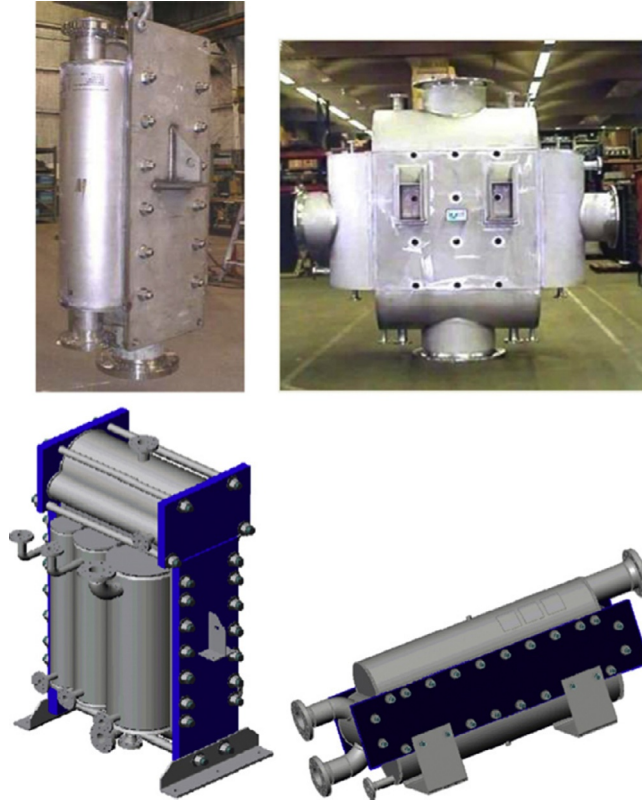


Figure 4.63 Welded plate-and-frame hybrid design.

thickness, determine the pressure rating of the coil. The high-pressure coil should be rated for the Shut-in-tubing-pressure (SITP). The low-pressure coil should be determined from the maximum allowable working pressure of the downstream equipment.

4.6.7.2 Calculate the Heat Duty (q)

The heat duty (q) for each coil should be calculated using the following equation:

$$\text{MTD} = \frac{\Delta t_1 - \Delta t_2}{l_{N_e} \left(\frac{\Delta t_1}{\Delta t_2} \right)} \quad (4.7)$$

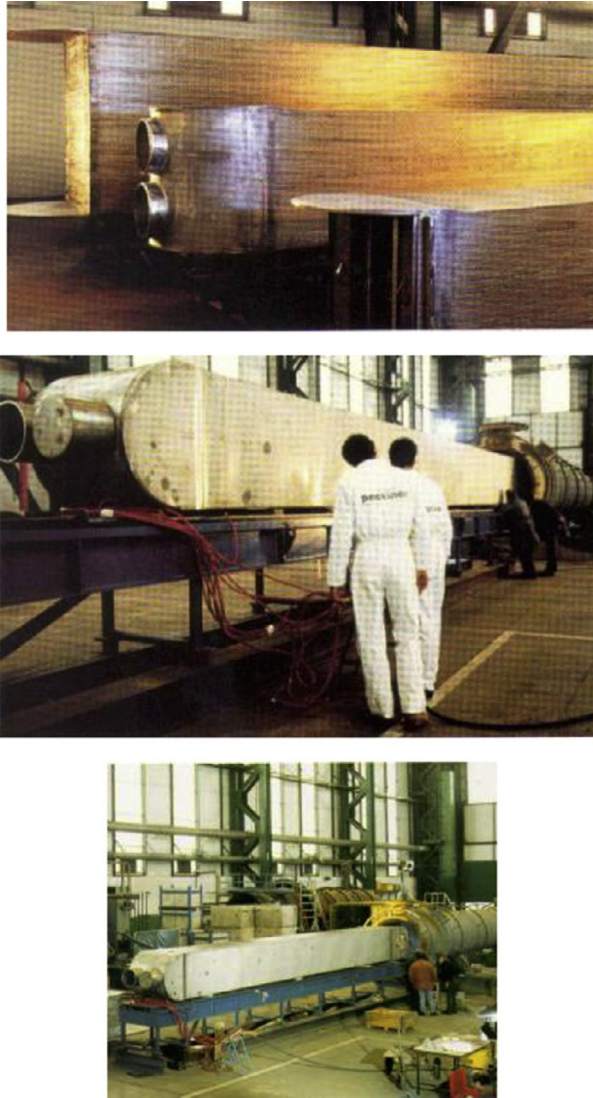


Figure 4.64 Welded plate-and-frame exchangers.

where Δt_1 = temperature difference between coil inlet and bath;
 Δt_2 = temperature difference between coil outlet and bath.

4.6.7.3 Calculate the Overall Heat Transfer Coefficient (U)

Calculate U using the equations presented in [Chapter 3](#) or assume U using the appropriate table or chart (typical range 44–53 Btu/h ft² °F).

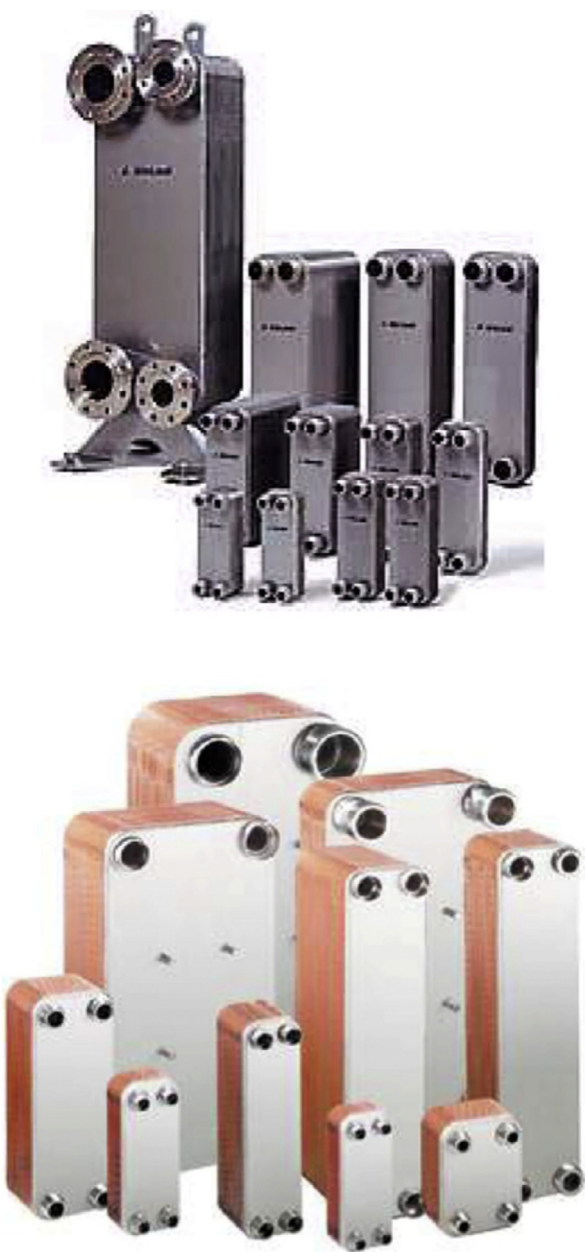


Figure 4.65 Brazed plate-and-frame exchanger design.

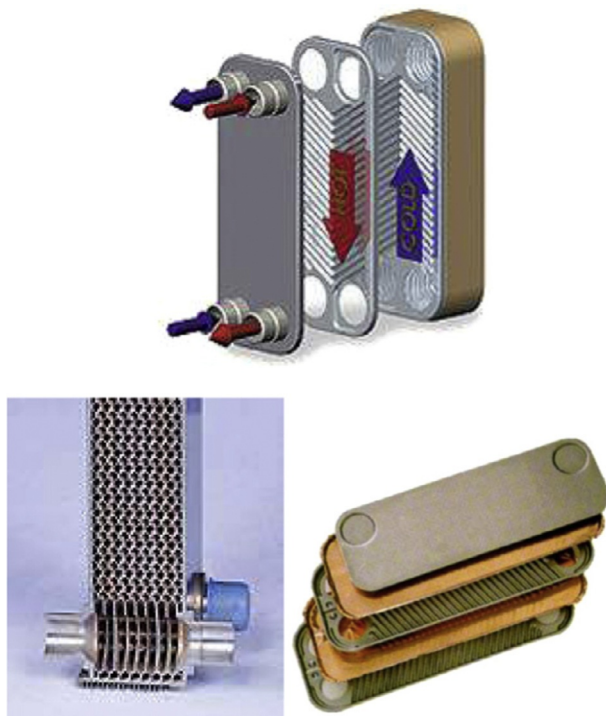


Figure 4.66 Brazer plate-and-frame exchanger design.



Figure 4.67 Compact doesn't necessarily mean small! Welded plate-and-frame exchanger (Alfa Laval/Packinox).

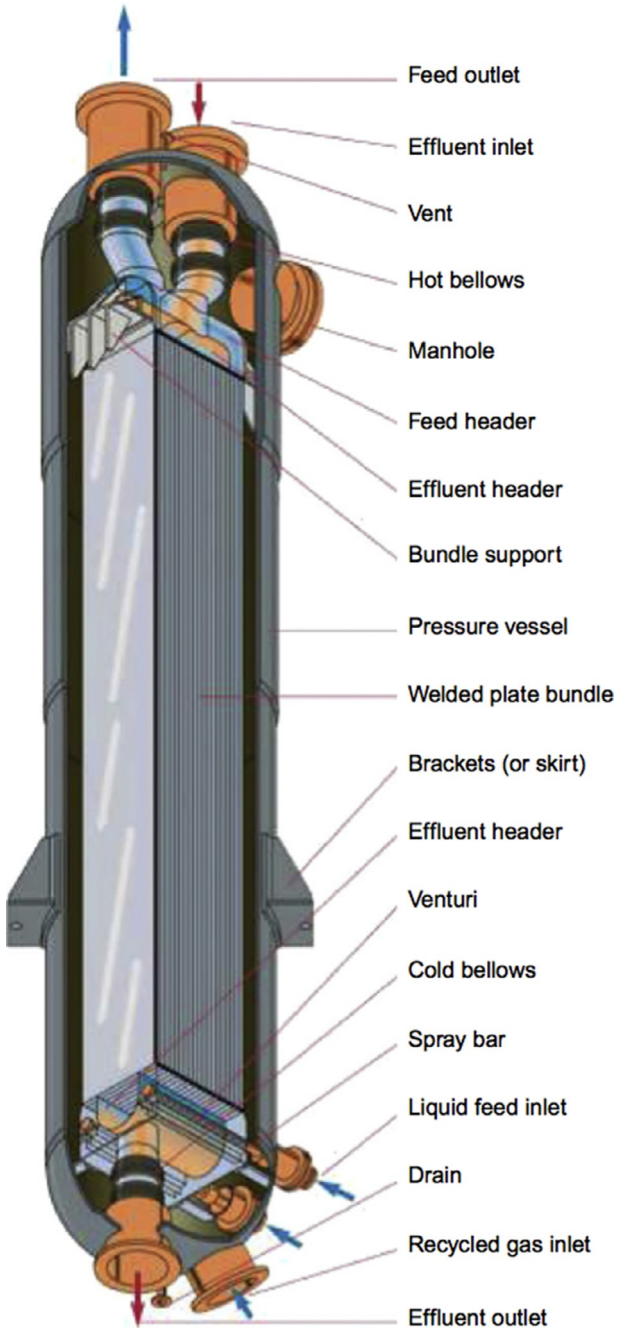


Figure 4.68 Welded plate-and-frame exchanger (Alfa Laval/Packinox).

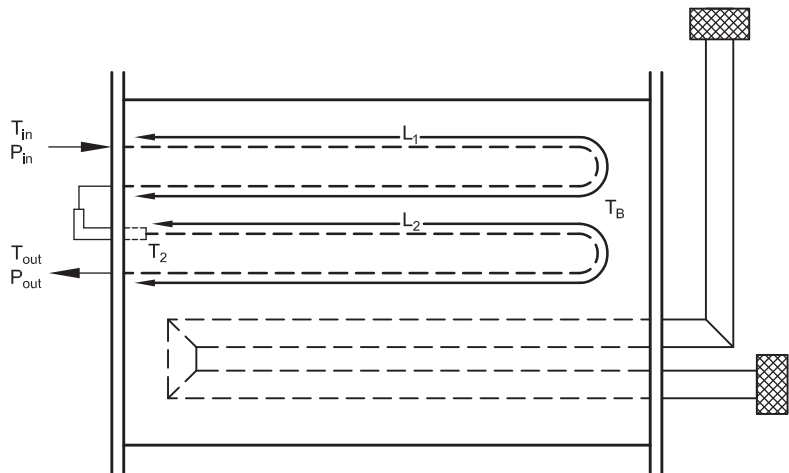


Figure 4.69 Schematic of a line heater.

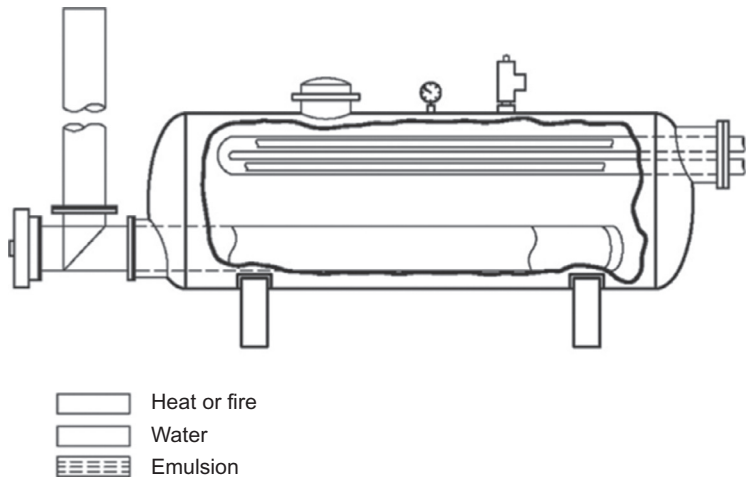


Figure 4.70 Cutaway of an indirect-fired heater.

4.6.7.4 Calculate Coil Length

Because U , MTD, q , and the diameter of the pipe are known, the length of the coil can be calculated from the following equation:

$$L = \frac{12q}{\pi(\text{MTD})Ud} \tag{4.8}$$

where d = coil diameter, in.

Equation (4.8) describes an overall length required for the coil.

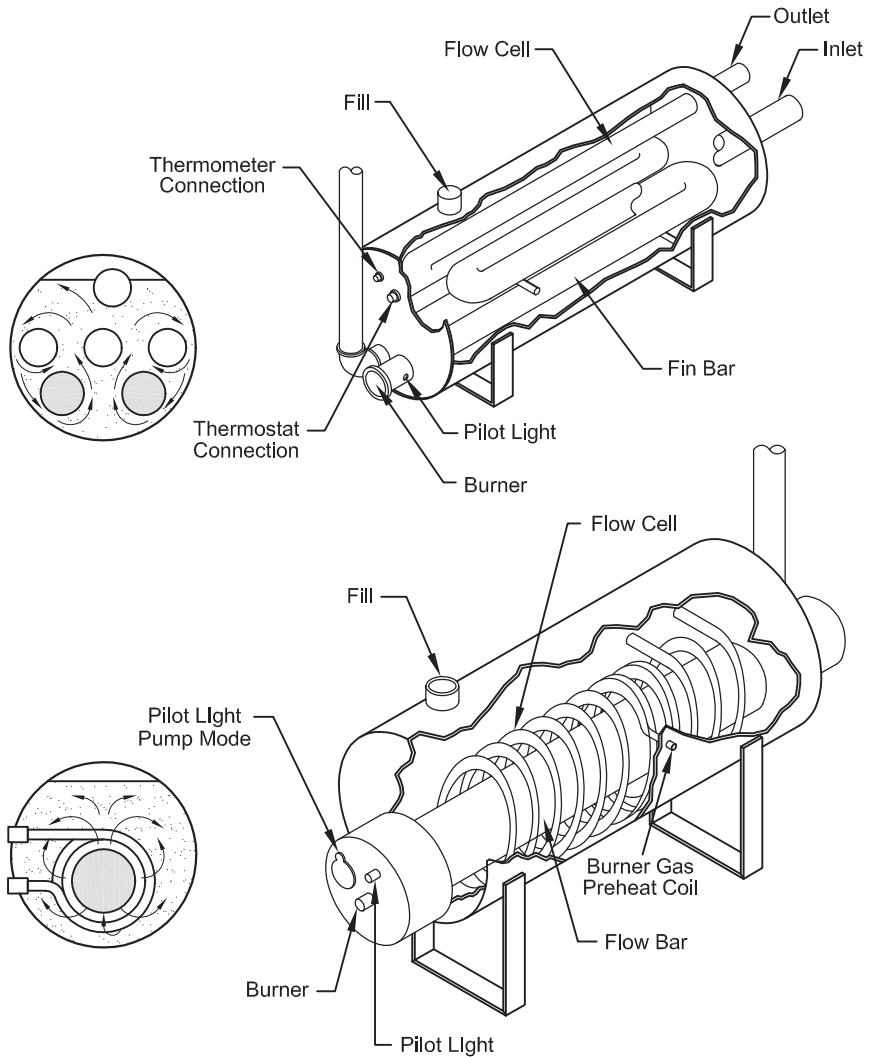


Figure 4.71 Types of indirect heaters.

4.6.8 Heater Sizing

Based on previous information, choose the shell length and number of passes. As the shell length decreases, the number of passes increases and a larger diameter is required. The preceding procedure is helpful for reviewing existing designs or vendor proposals. However, it is more economically advantageous to select standard size heaters from the manufacturer. See [Figure 4.73](#).

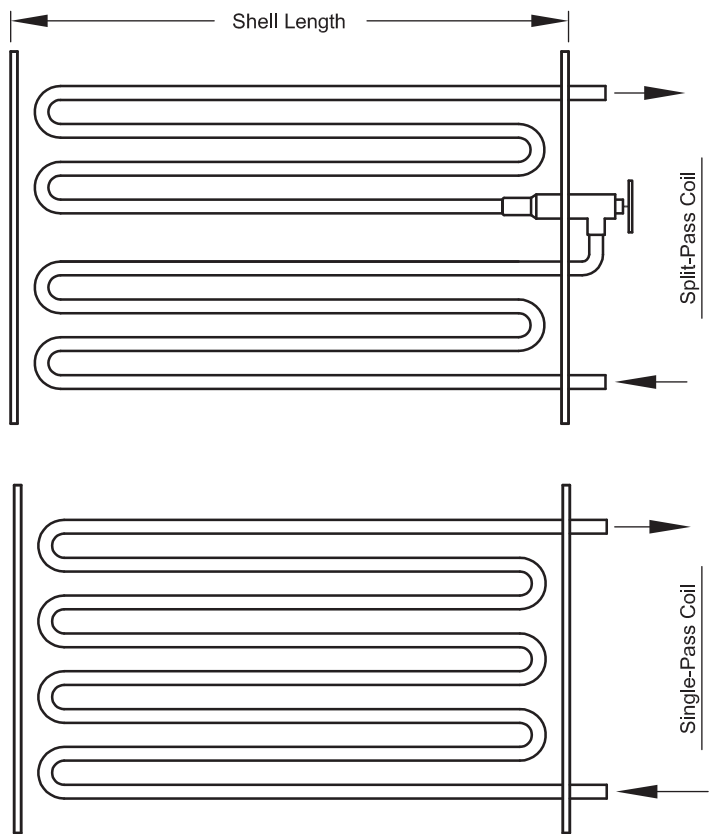
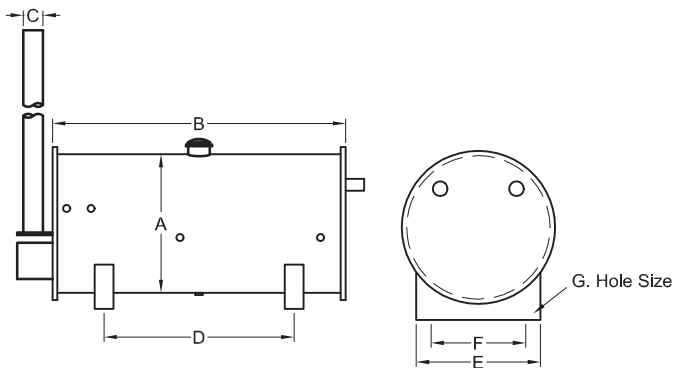


Figure 4.72 Indirect heating coils.

Table 4.7 Standard Size Fire Tubes for Indirect-Fired Heaters

250,000 Btu/h	4,000,000 Btu/h
500,000 Btu/h	4,500,000 Btu/h
750,000 Btu/h	5,000,000 Btu/h
1,000,000 Btu/h	6,000,000 Btu/h
1,500,000 Btu/h	7,000,000 Btu/h
2,000,000 Btu/h	8,000,000 Btu/h
2,500,000 Btu/h	9,000,000 Btu/h
3,000,000 Btu/h	10,000,000 Btu/h
3,500,000 Btu/h	



NOMINAL DIMENSIONAL DATA

Heater	A	B	C	D	E	F	G
BTU HR	Ft. In.	Ft. In.	Ft. In.	Ft. In.	Ft. In.	Ft. In.	In. only
250,000	2'-0"	7'-6"	0'-8"	5'-6"	1'-0"	1'-5"	3/4"
500,000	2'-6"	10'-0"	0'-10"	6'-0"	1'-9"	1'-5"	11/16"
750,000	3'-0"	12'-0"	0'-12"	6'-0"	2'-2"	1'-10"	11/16"
1,000,000	3'-6"	14'-4"	1'-2"	11'-0"	3'-0"	2'-4"	3/4"
1,500,000	4'-0"	17'-6"	1'-4"	12'-6"	3'-6"	3'-0"	3/4"
2,000,000	5'-0"	20'-0"	1'-8"	12'-6"	4'-4"	3'-0"	7/8"

SPECIFICATIONS

Heater Furnace Input BTU/HR	Shell Size O.D. x Lgt.	Std. No. & Size Tubes	Coil W.P. PSI	Std. Mean Coil Area Sq. Ft.	Approx. Coil. Lin. Ft.	Water Fill Vol: Bbls.	Shipping Weight Pounds
250,000	24"x7' 6"	8-2"XH	3372	29.5	54	2.9	1,400
250,000	24"x7' 6"	8-2"XXH	6747	26.5	54	2.9	1,610
500,000	30"x10' 0"	8-2"XH	3372	42.6	76	6.0	2,210
500,000	30"x10' 0"	8-2"XXH	6747	38.3	76	6.0	2,510
750,000	36"x12' 0"	10-2"XH	3372	64.4	114	10.5	2,875
750,000	36"x12' 0"	10-2"XXH	6747	58.8	114	10.5	3,325
750,000	36"x12' 0"	6-3"XH	3150	59.4	70.9	10.3	3,030
750,000	36"x12' 0"	6-3"XXH	6300	58.8	70.9	10.3	3,615
1,000,000	42"x14' 4"	12-2"XH	3372	93.4	166	17.9	4,060
1,000,000	42"x14' 4"	12-2"XXH	6747	85.9	166	17.9	4,725
1,000,000	42"x14' 4"	8-3"XH	3150	94.8	113.2	17.5	4,390
1,000,000	42"x14' 4"	8-3"XXH	6300	85.9	113.2	17.5	5,335
1,500,000	48"x17' 6"	14-2"XH	3372	134.0	237	28.7	5,650
1,500,000	48"x17' 6"	14-2"XXH	6747	120.5	237	28.7	6,600
1,500,000	48"x17' 6"	10-3"XH	3150	145.0	173.1	28.0	6,235
1,500,000	48"x17' 6"	10-3"XXH	6300	131.4	173.1	28.0	7,675
2,000,000	60"x20' 0"	16-2"XH	3372	175.7	311	51.8	10,110
2,000,000	60"x20' 0"	16-2"XXH	6747	158.0	311	51.8	11,360
2,000,000	60"x20' 0"	10-3"XH	3150	165.9	198.1	51.2	10,580
2,000,000	60"x20' 0"	10-3"XXH	6300	150.4	198.1	51.2	12,240

Figure 4.73 Standard indirect heaters and coil sizes.

4.7 DIRECT-FIRED HEATERS

4.7.1 Overview

Direct-fired heaters are used when large amounts of heat input are required. They consist of two types:

- Those that heat the fluid with a fire tube directly
- Those that heat the fluid using both radiant and convection

As shown in [Figure 4.74](#), direct-fired heaters use a fire tube to heat the fluid directly. Oil passes through an inlet distributor and is heated directly by a fire box. They are quick, efficient (75–90%), and the initial cost is relatively low. If fuel gas is available, utilization of a direct-fired heater for oil treating (especially high-volume oil treating) should be considered.

They are hazardous and require special safety equipment. Scale may form on the oil side of the fire tube, preventing the transfer of heat from the fire box to the oil emulsion. Heat collects in the steel walls under the scale. Metal softens, buckles ([Figures 4.75](#) and [4.76](#)), and eventually ruptures, allowing oil to flow into the fire box. The resultant blaze, if not extinguished, will be fed by the incoming oil stream.

Direct-fired heaters may also heat fluid using radiant and convection heating. Because the temperature in both sections is very high, an intermediate heat transfer fluid is typically used. There is a wide variety of heating configurations; choice depends on

- Fuel cost
- Thermal efficiency
- Temperature desired

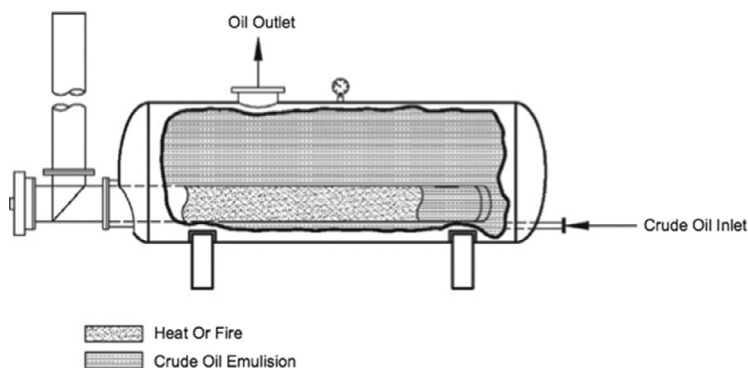


Figure 4.74 Cutaway of a direct-fired heater.



Figure 4.75 Fire tube scale prevents heat transfer.



Figure 4.76 Fire tube scale and potential hot spots.

- Size of the heat load
- Fluid being heated

Two basic configurations, include

- Horizontal tubes
- Vertical tubes

4.7.2 Horizontal Tubes (Refer to Figure 4.77)

4.7.2.1 Cabin

- Radiant section normally lines the walls with burners in the floor
- Economical, efficient, and most popular
- Normal duty range: 10-100 MMBtu/h

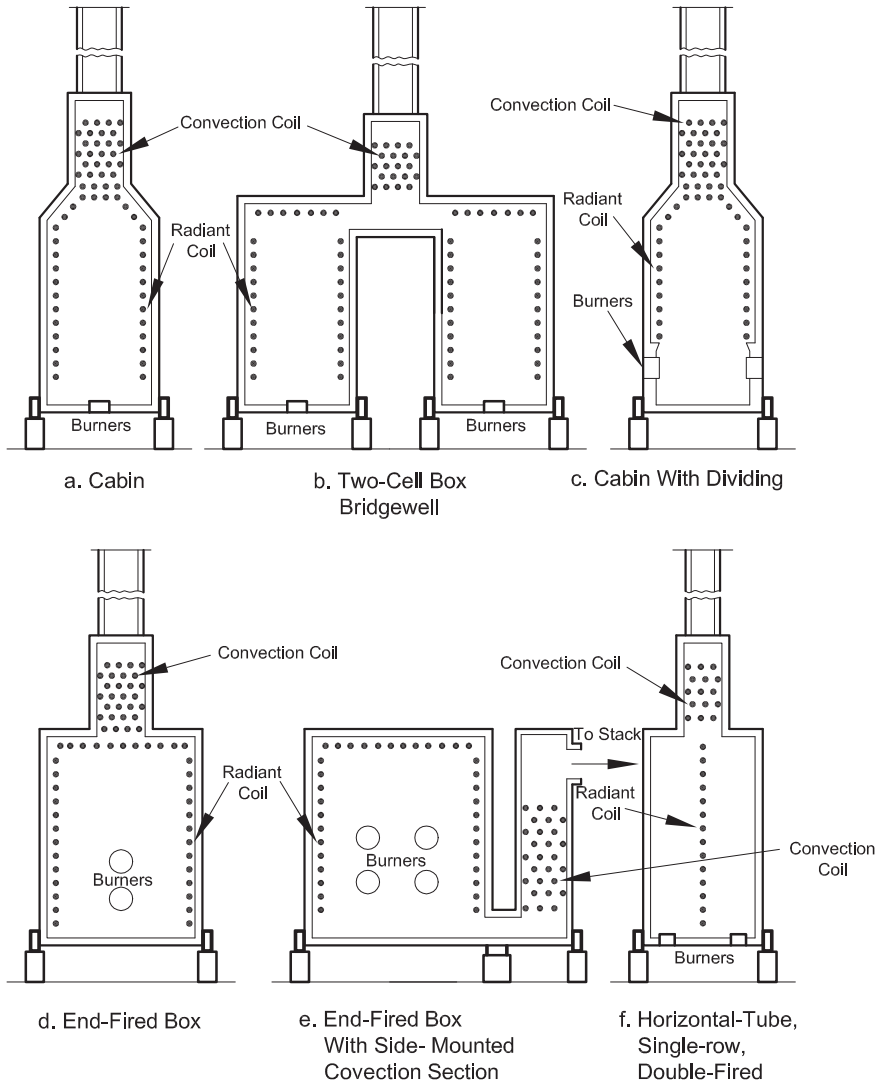


Figure 4.77 Basic types of direct-fired heaters with horizontal tube.

4.7.2.2 Two-Cell Box

- Only two cells are shown, but three or four can be used
- Vertically fired from floor for an economical, high-efficiency design
- Normal duty range: 100–250 MMBtu/h

4.7.2.3 Cabin with Bridge-Wall

- Provides two sections that can be fired individually
- Can be fired horizontally or vertically
- Normal duty range: 20–100 MMBtu/h

4.7.3 End Fired Box

- Horizontally fired as name implies
- Normal duty range: 5–50 MMBtu/h
- End fired box with side-mounted convection section
- Older type unit that may be used in new installations with high ash, poor grade fuels
- More expensive design
- Normal duty range: 50–200 MMbtu/h

4.7.3.1 Single Row, Double Fired

- Consists of one or more cells shown
- Used for reactor-fed heating units
- Normal duty range: 20–50 MMbtu/h

4.7.4 Vertical Tubes (Refer to [Figure 4.78](#))

4.7.4.1 All Radiant

- Low-cost, low-efficiency design that is compact
- Normal duty range: 0.5–20 MMbtu/h

4.7.4.2 Cylindrical, Helical Coil

- Basic low-cost, low-efficiency alternative to the “all radiant” type
- Not feasible to have parallel flow coils for fluid
- Normal duty range: 0.5–20 MMbtu/h

4.7.4.3 Cylindrical, with Cross-Flow Convection

- Popular new vertical flow units
- Economical, efficient, and compact
- Normal duty range: 10–200 MMbtu/h

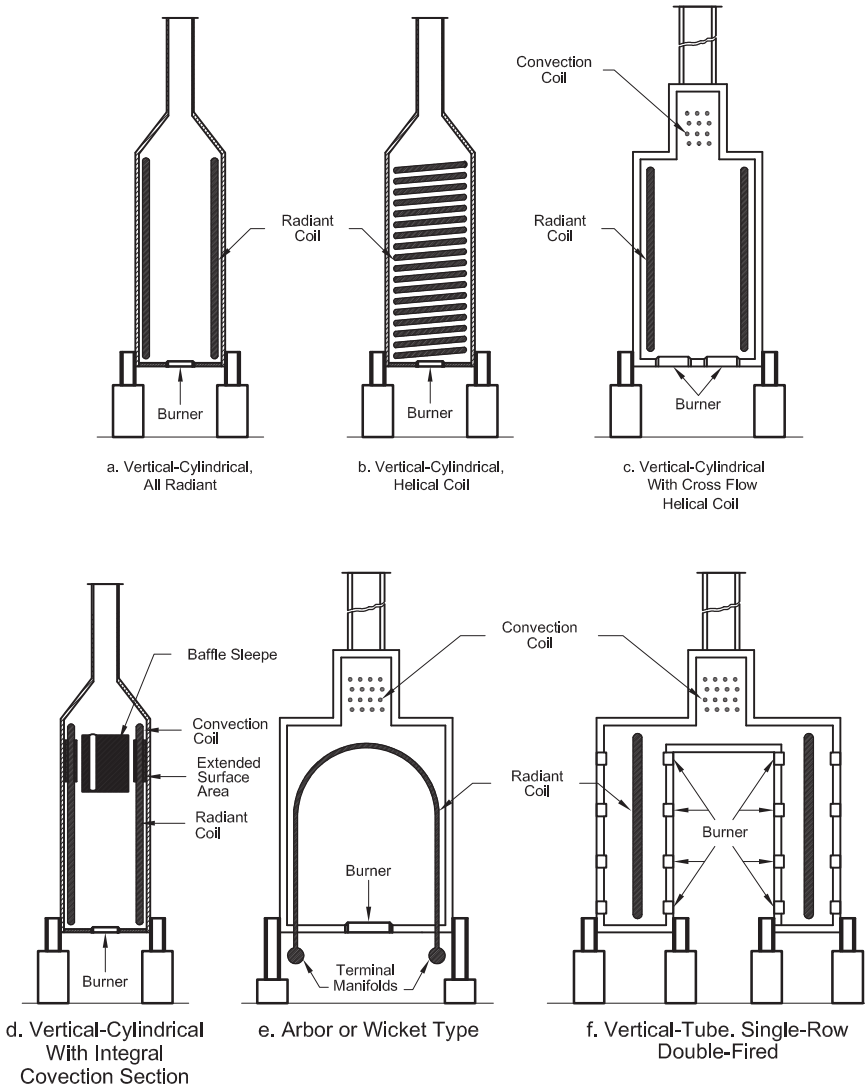


Figure 4.78 Basic types of direct-fired heaters with vertical tube heaters.

4.7.4.4 Cylindrical, with Integral Convection

- Many existing units in service
- Not commonly used due to its limited thermal efficiency
- Normal duty range: 0-100 MMbtu/h

4.7.4.5 Arbor or Wicket

- Used to heat large quantities of gas where low pressure drop is desired

- Several arbor coils may be used in one heater unit
- Normal duty range: 50–100 MMbtu/h

4.7.4.6 Single-Row, Double-Fired

- Most expensive configuration, but provides high and rather uniform heat flux
- Normal duty range: 20–125 MMbtu/h

4.7.5 Development of “Hot Spots” and Tube Failure

Result from

- High flame temperatures
- Low convection film coefficient
- Tube metallurgy selection must be evaluated in terms of initial cost and service life.
- Choice of material, method of welding, configuration used, etc., must be based on actual field experience.

4.7.6 Thermal Efficiency Considerations

Thermal efficiency considerations include:

- Control of excess air.
- Additional convection heat recovery (Figure 4.79).
- Preheating the combustion air stack.
- Using turbine exhaust gas (waste heat) (Figure 4.80). Excess air requires fuel to heat up ambient air to the stack temperature. Thermal efficiency is inversely proportional to excess air; stack air varies directly.
- Use of studded or fin type convection tubes offers enhanced heat recovery from the hot combustion gases (Figure 4.81).
- Use of extended surface on the low coefficient side tends to reduce fouling inside the tubes, which may be a major problem. However, said surfaces foul easier.
- Typically combustion air is used to preheat the combustion gas. Thermal efficiency may reach 90% with the stack temperature lowered to 300–350 °F. This tends to elevate combustion temperatures, which increases radiant heat transmission and tube surface temperature.

Gas turbine exhaust can be a super substitute for air because

- It contains 17–18% oxygen
- Is already preheated to 800–900 °F
- Combustion gas is flowed through a convection section to heat process streams

This heat is used directly or indirectly to generate steam, run boilers, regenerate solid desiccant units, heat gas and liquid streams, and so on.

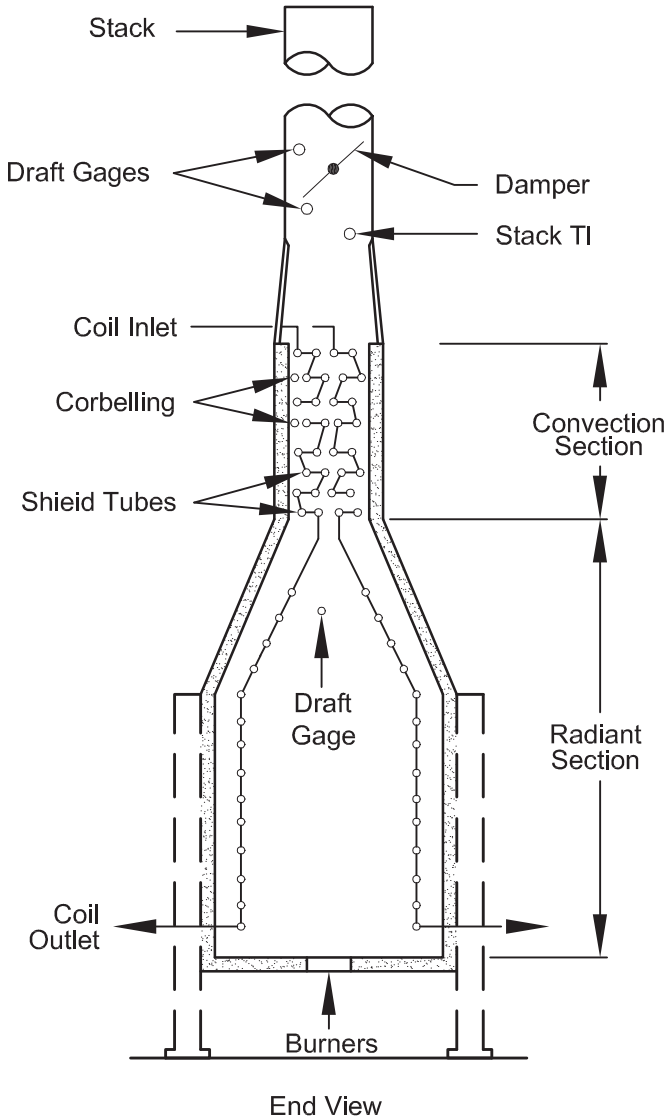


Figure 4.79 Fired heater.

4.7.7 Determining Required Heat Input

The general heat transfer equation is expressed by the following equation:

$$q = WC_p \Delta T \quad (4.9)$$

where q = heat required, Btu/h; W = flow rate, lb/h; C_p = specific heat constant, Btu/lb °F, =1.0 for water, =0.5 for oil (average); ΔT = differential temperature, °F.

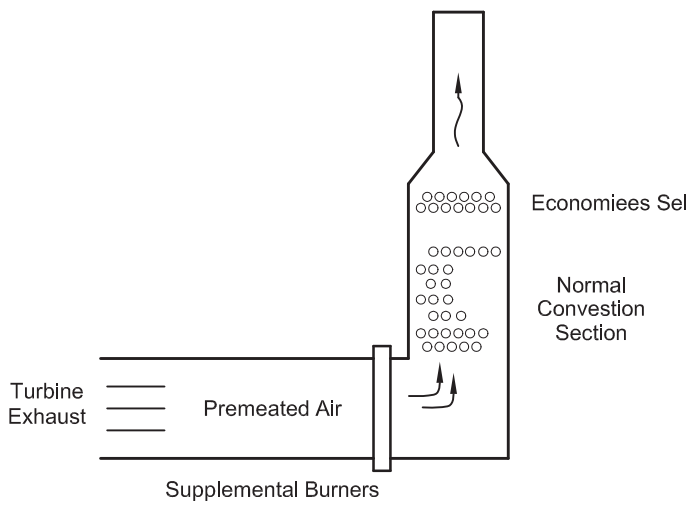


Figure 4.80 Waste heat recovery unit (exhaust heated).

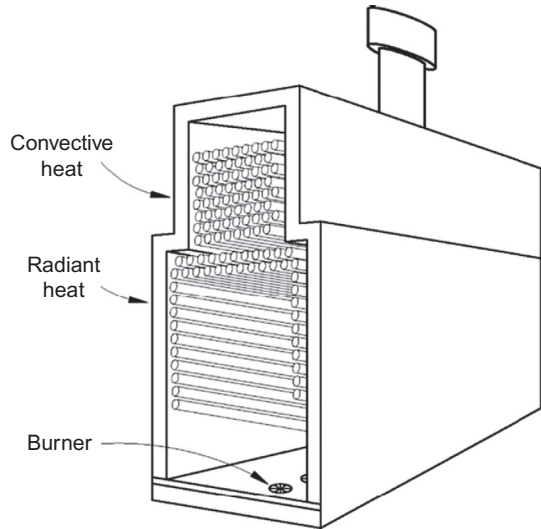


Figure 4.81 Direct-fired furnace.

4.7.7.1 Flow Rate Determination

Because water weighs 350 lbs/bbl, the flow rate can be expressed as follows, employing conversion factors:

$$W = \left(\frac{350}{24} \right) (SG)_L Q_L \tag{4.10}$$

where $(SG)_L$ = specific gravity of liquid; Q_L = liquid flow rate, BPD.

4.7.7.2 Total Heat Required

Total heat required is expressed by the following equation:

$$q = q_0 + q_w + q_g + q_{\text{lost}} \quad (4.11)$$

where q_0 = heat required to heat the oil, $= [(350/24)(SG)_o Q_o](0.5) \Delta T$;
 q_w = heat required to heat the water, $= [(350/24)(SG)_w Q_o](1.0) \Delta T$; $q_g = 0$.

Substituting into Equation (4.11)

$$q = \frac{350}{24} [(SG)_o Q_o(0.5) + (SG)_w Q_w(1)] \Delta T + q_{\text{lost}} \quad (4.12)$$

Assuming that the free water has been removed so that

$Q_w = 0.1 Q_o$, $(SG)_w = 1.0$, and $q_{\text{lost}} = (0.1)(q)$, then the following equation exists:

$$q = 15 [(0.5)(SG)_o Q_o + (SG)_w Q_w] \Delta T + q_{\text{lost}} \quad (4.13)$$



4.8 AIR-COOLED EXCHANGERS

4.8.1 General Considerations

Air-cooled exchangers are expensive but are also cost-efficient. They are usually designed by the manufacturer. Air cooling is preferred if the ambient air temperature is low enough to provide efficient cooling. Cooling towers have no application in offshore, arctic, or humid locations. They are popular where water for cooling is available because they are

- Mechanically simple
- Flexible
- Able to eliminate the nuisance and cost of water treating

In warm climates

- They may not be capable of producing as low a temperature as water
- Study should be conducted for cooling alternatives

Common uses

- To cool hot fluid to something near ambient
- As an interstage cooler in compression

4.8.2 Typical Air-Cooled Exchanger Configurations

- Four rows of tubes on staggered pitch for moderate temperature ranges
- Six rows of tubes for a large temperature range service
- Pitch of 2½ inch (63.5 mm) with equilateral triangular layout 1-in. OD tubes with ten 5/8-in. (15.875-mm) high aluminum fins per inch

- 20:1 finned to outside bare surface area ratio
- 30-ft (9144 mm) tube length (when mounted over pipe way)
- At least two fans per bay (half of which are auto-variable pitch)
- Aluminum or plastic fan blades
- Draft types
- Forced draft
 - Less expensive
 - Tube bundle located on discharge of fan
 - Used when outlet air is too hot for fan parts
- Induced draft
 - More efficient
 - Tube bundle located on suction of fan
 - Used when fans off-performance is needed

Multiple services (bundles) located in same bay

- Caution of “overcooling”

Winterization

- Warm air recirculation with steam

4.8.3 Advantages of Forced Draft Design

- Easy to remove and replace bundles
- Easier to mount motors or other drivers with short shafts
- Accessible lubrication, maintenance, and so on
- Simplified shipping and mounting due to reinforced straight side panels that form a rectangular box type of plenum, permitting complete pre-assembled and shop-tested units
- Best adapted for cold climate operation with warm air recirculation
- Requires less horsepower for an air temperature rise $>50^{\circ}\text{F}$

4.8.4 Disadvantages of Forced Draft Design

- Possibility that hot air leaving the top will flow around the unit and be drawn through again
- Less efficient air distribution over the bundle
- Harder to clean when covered with lint, bugs, or debris
- Hoods do not offer protection from weather

4.8.5 Advantages of Induced Draft Design

- Easier to shop assemble, ship, and install
- Hoods offer protection from weather

- Easier to clean underside when covered with lint, bugs, or debris
- More efficient air distribution over the bundle
- Less likely to be affected by hot air recirculation
- Requires less horsepower for an air temperature rise $<50^{\circ}\text{F}$

4.8.6 Disadvantages of Induced Draft Design

- More difficult to remove bundles for maintenance
- High temperature service is limited due to the effect of hot air on fans
- More difficult to work on the fan assembly and adjust blades due to heat from bundle and their location
- Tube lengths
 - 6–50 ft
- Tube diameter
 - 5/8–1 in.
 - Air is nonfouling and offers low heat transfer efficiency
- Fins
 - 1/2–1-in. height
 - 7–11 per inch
- Depth
 - 3–8 rows of fin tubes
 - Triangular pitch with fins separated by 1/16–1/4 in.
- Bay widths
 - 4–30 ft
- Fan diameters
 - 3–16 ft

See [Figures 4.82–4.91](#).

4.8.7 Air-Side Control

Adjustment must be made to ensure adequate cooling while not overcooling the process fluid.

- Process flow rate/heat duties change
- Air temperatures change
 - Season to season
 - Night to day

Too cool a gas temperature could lead to hydrate formation. In addition, too cool a lube oil temperature leads to high viscosity, which results in high Δp and poor lubrication. Process outlet temperature is controlled by:

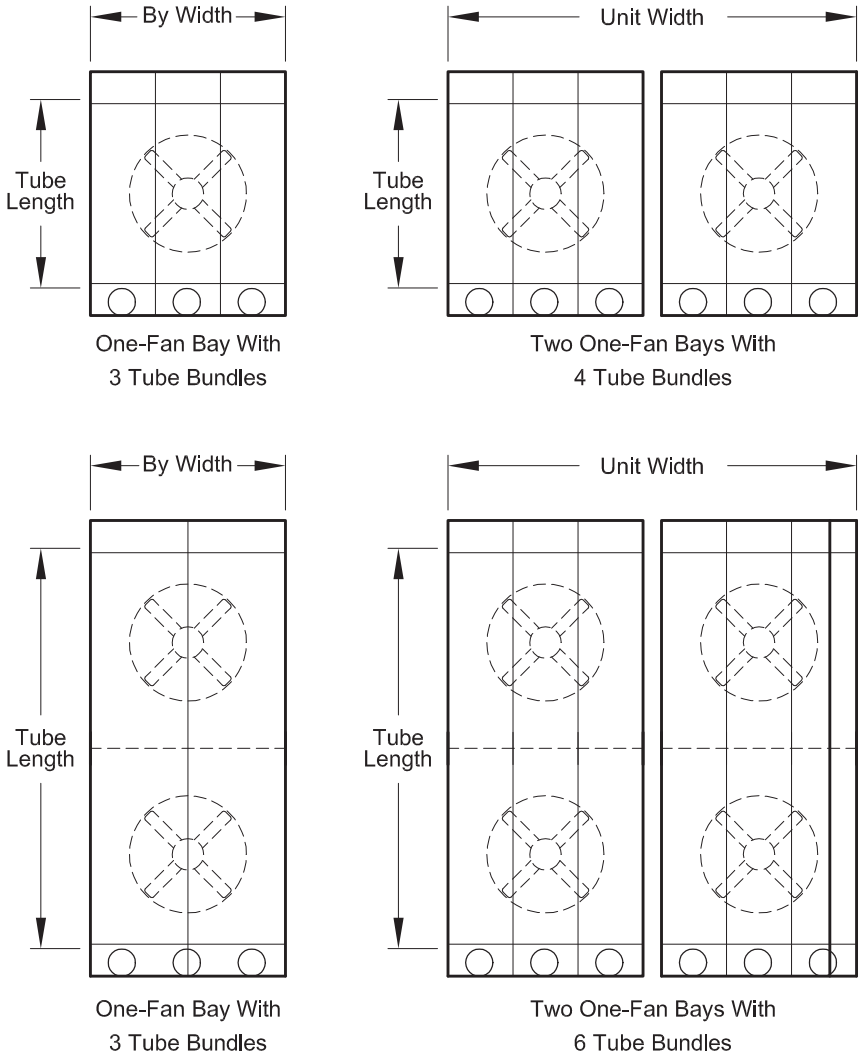


Figure 4.82 Air-cooled exchanger bundles and bays.

- Louvers (most common)
 - Energy inefficient
- Mechanics (seasonal or night/day temperature changes)
- Automatic sensors (senses process temperature)
- Variable pitch fan blades (second most common)
- Variable speed drivers (third most common)

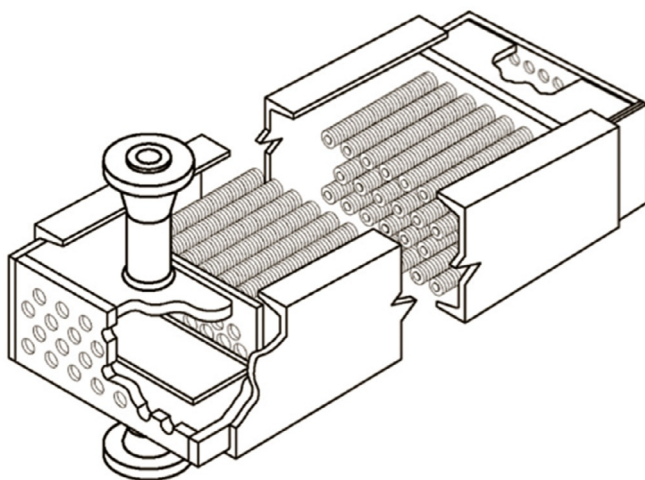
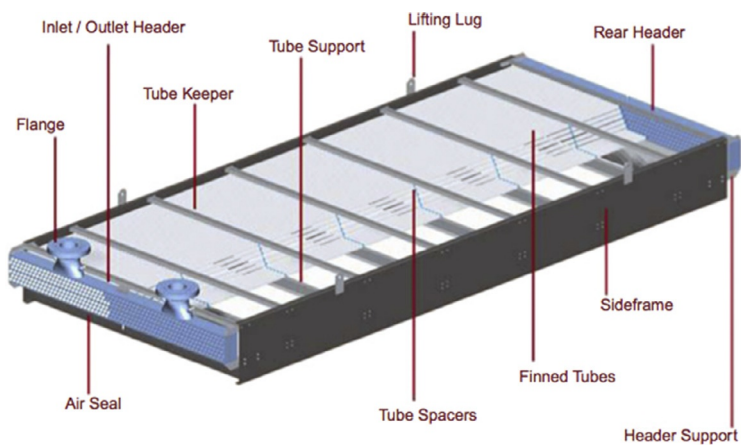
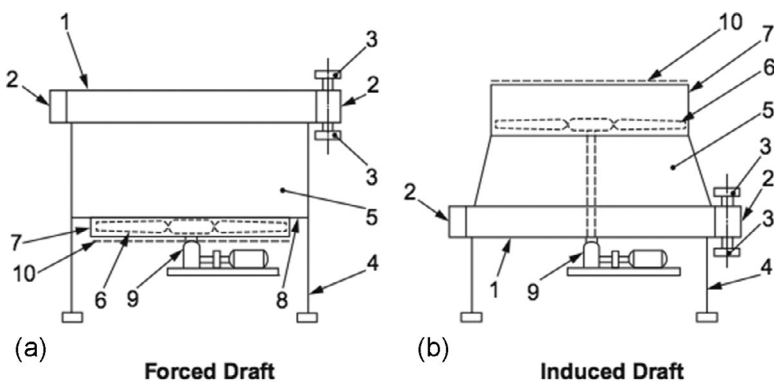


Figure 4.83 Air-cooled exchanger construction.



Key

- | | |
|---------------------|------------------|
| 1 tube bundle | 6 fan |
| 2 header | 7 fan ring |
| 3 nozzle | 8 fan deck |
| 4 supporting column | 9 drive assembly |
| 5 plenum | 10 fan guard |

Figure 4.84 Air-cooled exchanger plug header.

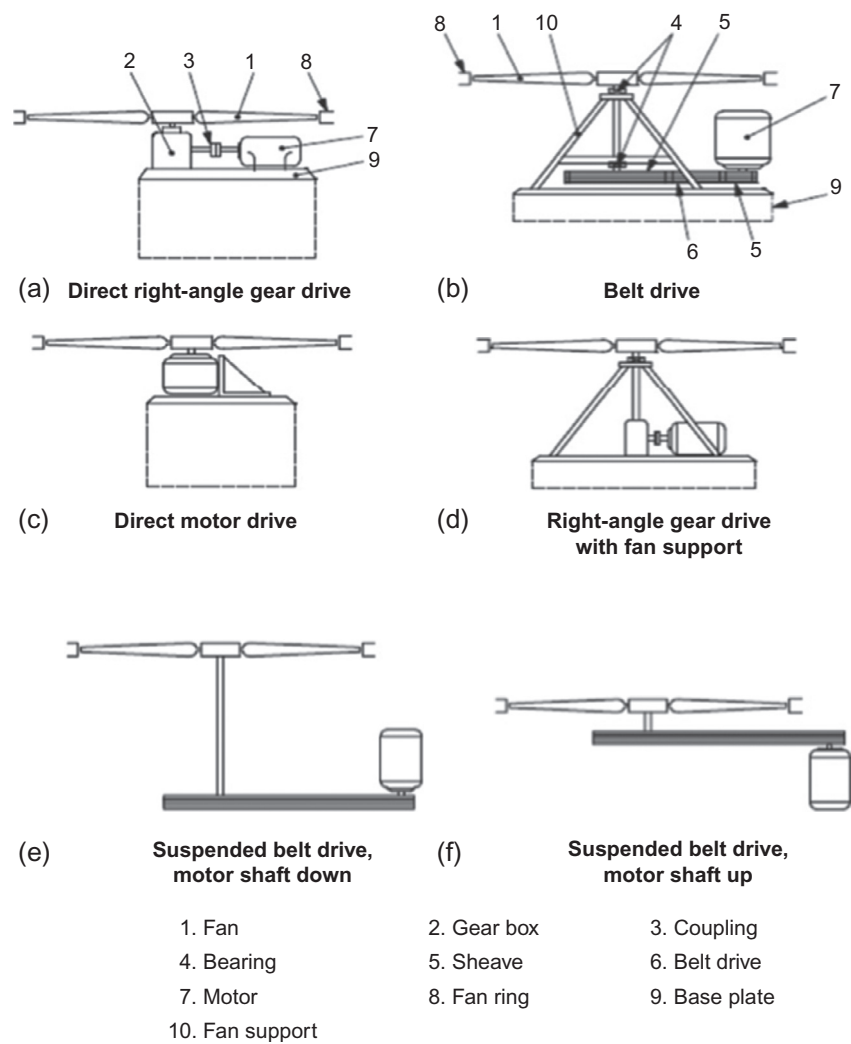


Figure 4.85 Air-cooled exchanger drives.

4.8.8 Procedure for Calculating Number of Tubes Required for Aerial Coolers

The procedure for calculating tubes required for aerial coolers is similar to sizing a shell-and-tube heat exchanger. Figure 4.92 lists approximate “U” values. Use “U_x” when extended surface area, including fins, is used for the area term in

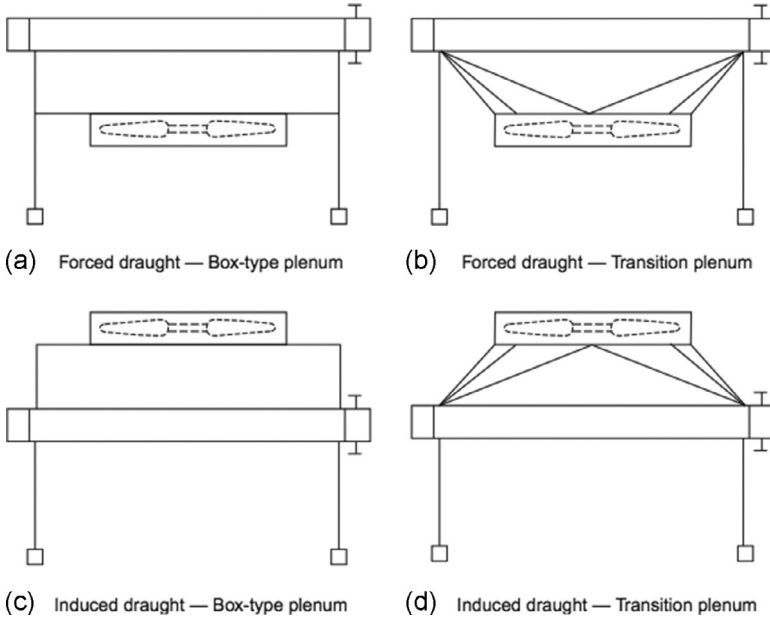


Figure 4.86 Air-cooled exchanger plenums.



Figure 4.87 Air-cooled exchanger fans.

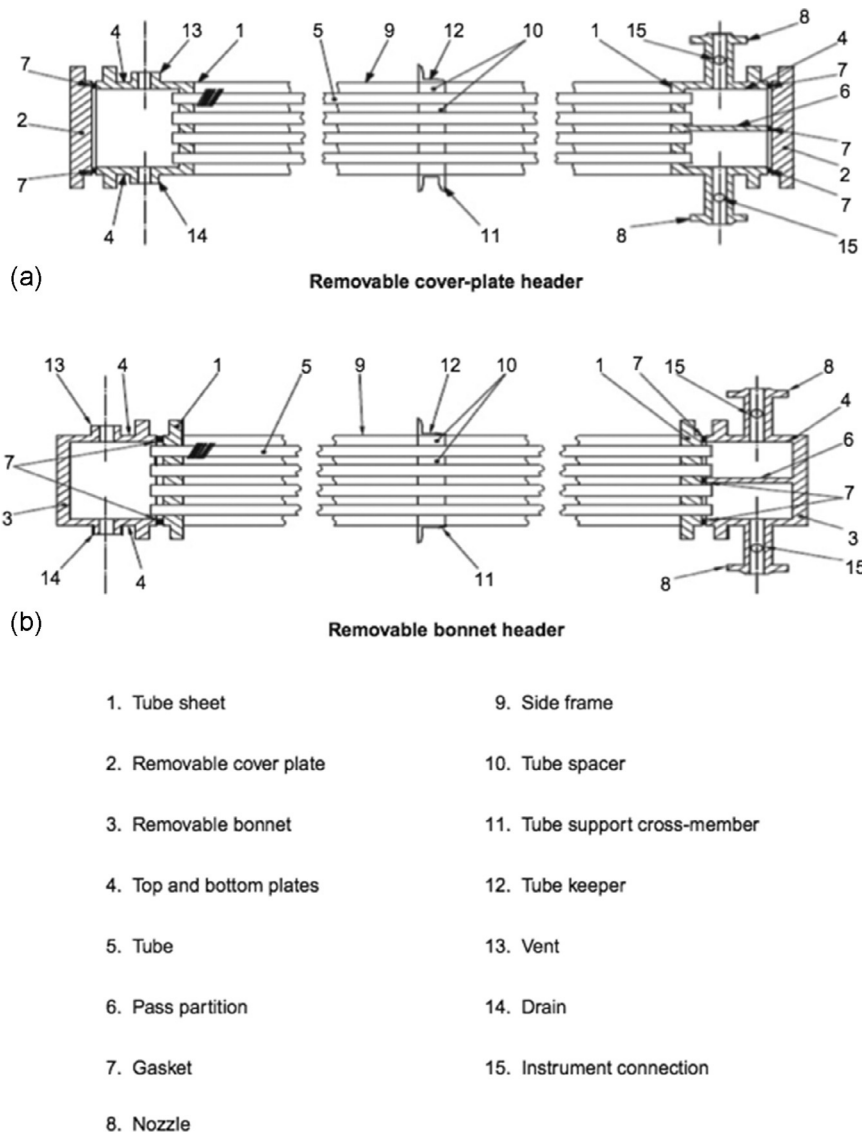
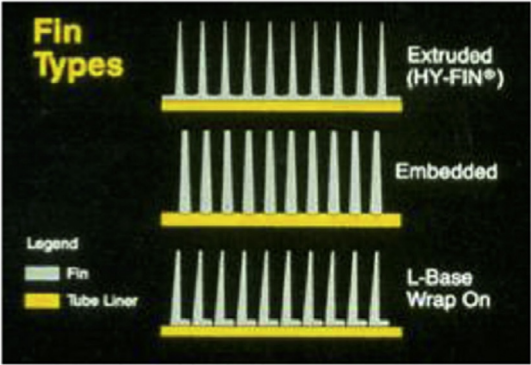
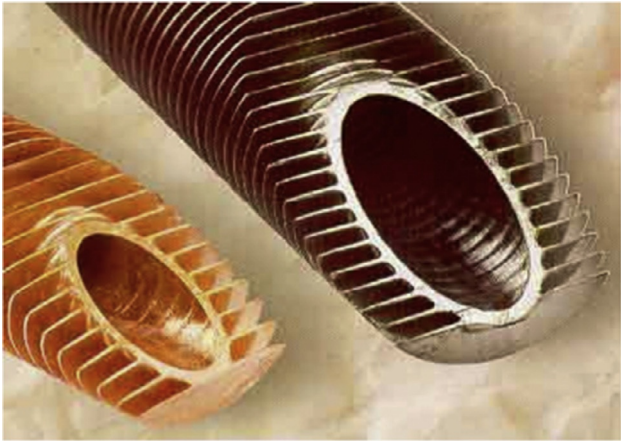


Figure 4.88 Air-cooled exchanger cover-plate header.

the general heat transfer equation. Use “ U_b ” when the outside surface area of the bare tube (neglecting fins) is used in the general heat transfer equation.

4.8.8.1 Procedure for Sizing Air-Cooled Exchanger

- (1) Calculate or use “ U_x ” from [Figure 4.92](#).
- (2) Approximate air temperature rise (exit air temperature)



Cutaway view of fin tubes



Figure 4.89 Cutaway of finned tubes.



Figure 4.90 Air-cooled exchangers.



Figure 4.91 Air-cooled exchanger.

Service	Finetube			
	½ in. By 9	5/8 in. By 10		
Water & water solutions				
	U_b	U_x	U_b	U_x
Engine jacket water (rf = 0.001)	110	7.5	130	6.1
Process water (rf = 0.002)	95	6.5	110	5.2
50-50 ethyl glycol-water (rf = 0.001)	90	6.2	105	4.9
50-50 ethyl glycol-water (rf = 0.002)	80	5.5	95	4.4
Hydrocarbon liquid coolers				
Viscosity C_p	U_b	U_x	U_b	U_x
0.2	85	5.9	100	4.7
0.5	75	5.2	90	4.2
0.2	65	4.5	75	3.5
0.2	45	3.1	55	2.6
0.2	30	2.1	35	1.6
0.2	20	1.4	25	1.2
0.2	10	0.7	13	0.6
Hydrocarbon gas coolers				
Temperature, °F	U_b	U_x	U_b	U_x
50	30	2.1	35	1.6
100	35	2.4	40	1.9
300	45	3.1	55	2.6
500	55	3.8	65	3.0
750	65	4.5	75	3.5
1000	75	5.2	90	4.2
Air and flue-gas coolers				
Use one-half of value given for hydrocarbon gas coolers				
Steam Condensers				
(Atmospheric pressure & above)				
	U_b	U_x	U_b	U_x
Pure steam (rf = 0.005)	125	8.6	145	6.8
Steam with non-condensables	60	4.1	70	3.3
HC condensers				
Pressure, psig	U_b	U_x	U_b	U_x
0° range	85	5.9	100	4.7
10° range	80	5.5	95	4.4
25° range	75	5.2	90	4.2
60° range	65	4.5	75	3.5
100° & over range	60	4.1	70	3.3
Other condensers				
	U_b	U_x	U_b	U_x
Ammonia	110	7.6	130	6.1
Freon 12	65	4.5	75	3.5

Note: U_b is overall rate based on bare tube area and U_x is overall rate based on extended surface.
Source: Gas Processor Suppliers Association, Engineering Data Book, 9th Edition.

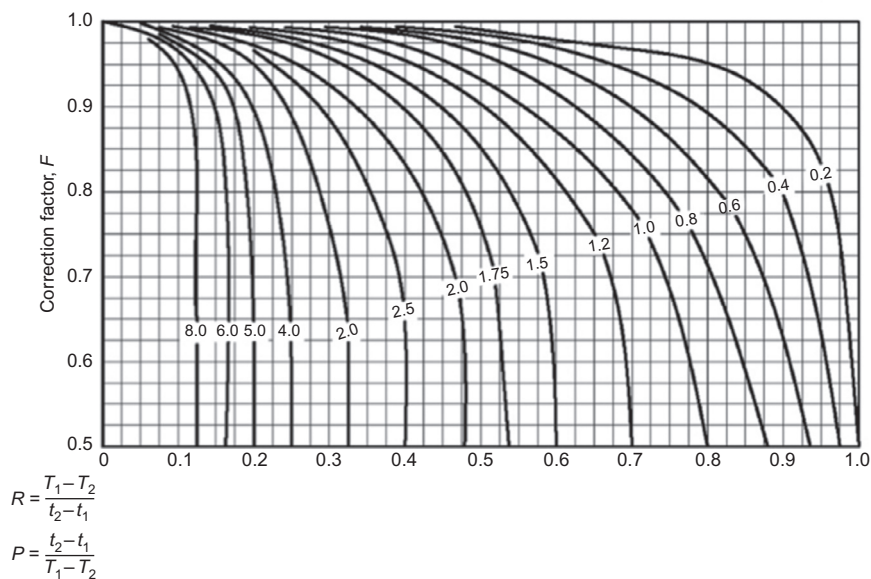
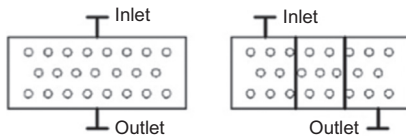
Figure 4.92 Typical heat-transfer coefficients for air coolers.

$$\Delta t_a = \left(\frac{U_x + 1}{10} \right) \left(\frac{t_1 + t_2}{2} - t_1 \right) \quad (4.14)$$

- (3) Calculate corrected LMTD using Figure 4.93.
- (4) Calculate heat transfer area required

Nomenclature:

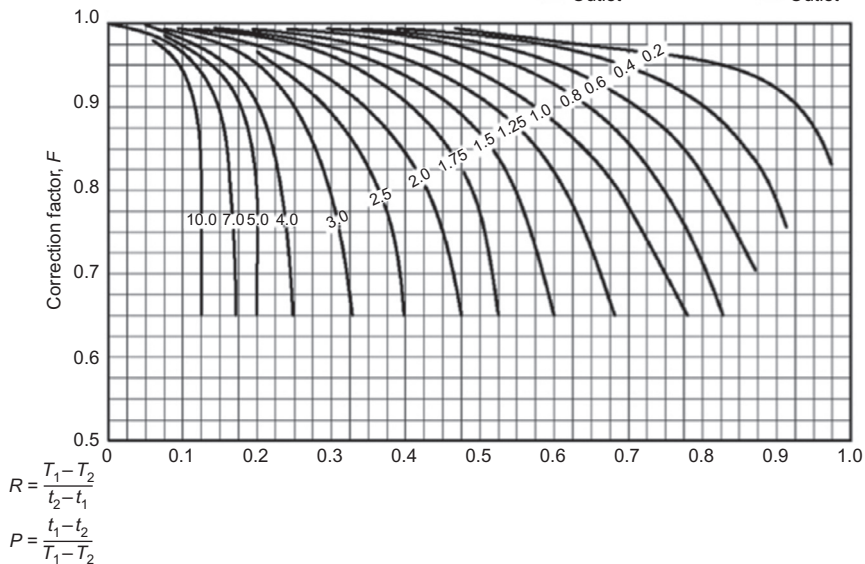
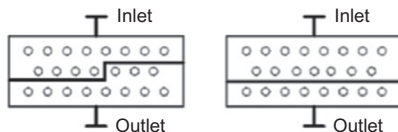
T_1 = Inlet temperature, tube
 T_2 = Outlet temperature, tube side
 t_1 = Inlet temperature, air side
 t_2 = Outlet temperature, air side



(a)

Nomenclature:

T_1 = Inlet temperature, tube
 T_2 = Outlet temperature, tube side
 t_1 = Inlet temperature, air side
 t_2 = Outlet temperature, air side



(b)

Figure 4.93 (a) LMTD correction factors; both fluids unmixed. (b) LMTD correction factor, two-pass cross-flow; both fluids unmixed.

Table 4.8 External Area of Fin Tube per ft² of Bundle Surface Area (APSF) for 1-in. OD Tubes

½-in. Height by 9 Fins/in.		5/8-in. Height by 10 Fins/in.		
Tube pitch	2-in. Δ	2¼-in. Δ	2½-in. Δ	2½-in. Δ
Three rows	68.4	60.6	89.1	80.4
Four rows	91.2	80.8	118.8	107.2
Five rows	114.0	101.0	148.5	134.0
Six rows	136.8	121.2	178.2	160.8

$$A_x = \frac{Q}{U_x \text{LMTD}} \quad (4.15)$$

$$F_a = \frac{A}{\text{APSF}} \quad (4.16)$$

- (5) Calculate bundle face area from [Table 4.8](#).
 (6) Calculate width from assumed length of tubes

$$\text{Width} = \frac{F_a}{L} \quad (4.17)$$

- (7) Calculate number of tubes

$$N_t = \frac{A_x}{(\text{APSF})L} \quad (4.18)$$

Tables, graphs, and procedures are given in the GPSA Engineering Data Book to calculate pressure drop in tubes. The actual U_x to compare to assumed size, number, and horsepower of fans is required.

Air heat effectiveness compared to water is 2% without fins and 20% with fins, while with fins on the air side (ignoring fouling effect of water on exchanger) it is 50–70% effective, depending on the type of water used.

4.8.9 Considerations When Sizing an Aerial Cooler

- Cooling tube design
- Fan arrangement
- Number of transfer bays
- Control of cooled fluid temperature as the ambient air temperature changes
- Usual mechanical considerations
- Maintenance of fins

Design is a trial-and-error calculation and should not be undertaken by a novice.

See also [Figure 4.94](#).



Figure 4.94 Air-cooled exchanger with finned tubes clogged with debris.



4.9 COOLING TOWERS

A cooling tower is a specialized heat exchanger in which two fluids (air and water) are brought into direct contact with one another to affect the transfer of heat. The entering air is cooled by being drawn through water with fans. Towers may have

- Splash fill
- Engineered/organized fill
- No fill

Cooling towers are used where water is expensive to treat or is in short supply. Essentially, a cooling tower is an aerial cooler preceded by an evaporating section ([Figure 4.95](#)).

When air temperature is low enough the water may be shut off. The water rate may be decreased at intermediate temperatures. There is a high initial cost for cooling towers, but they may offer a total cost savings in some applications. When equipped with controls to vary fan horsepower, the output of such a unit offers flexibility at minimum operating cost. Water is cooled primarily through evaporation and humidifying the air. [Figure 4.96](#)

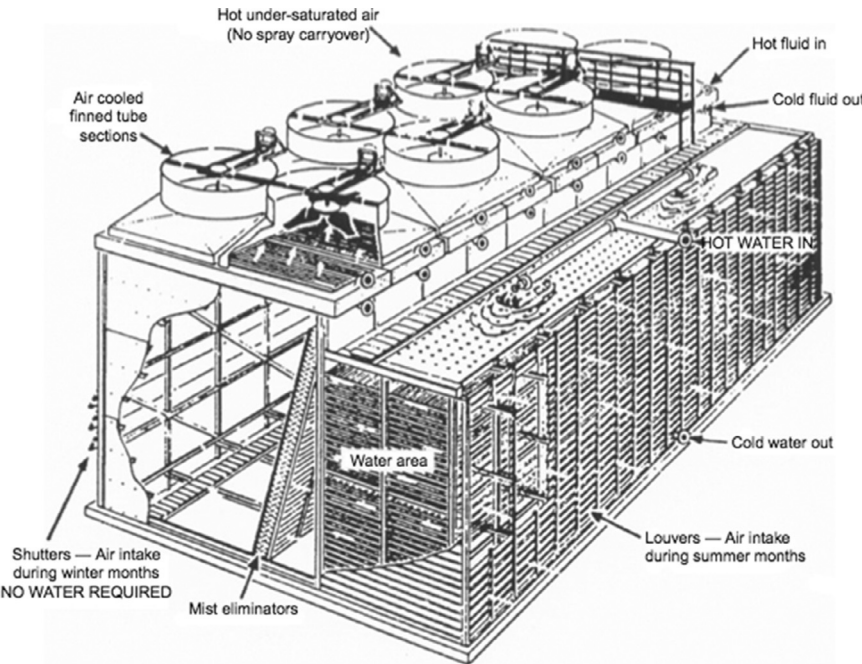


Figure 4.95 Combination air—water cooler.

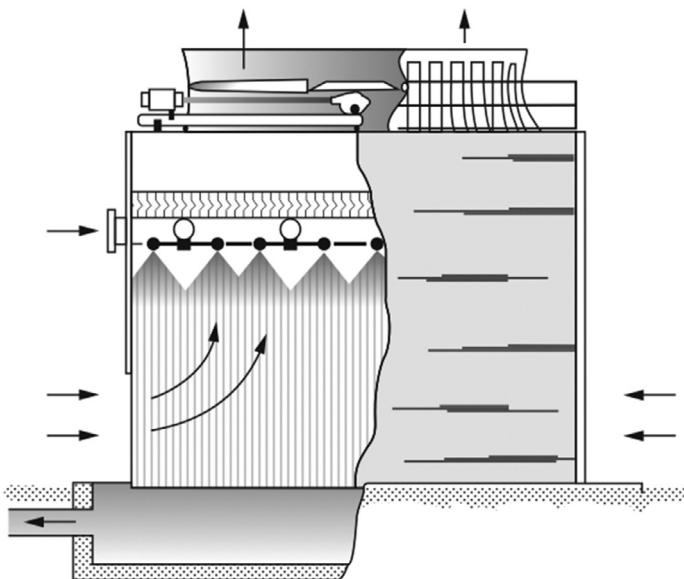


Figure 4.96 Spray-filled cooling tower.

shows a “spray-filled” tower. In this case, cooling is accomplished by spraying a flowing mass of water into a rain-like pattern through which an upward moving mass flow of cool air is induced by the action of a fan.



4.10 OTHER TYPES OF HEAT EXCHANGERS

4.10.1 Electric Heat Exchangers

Electric heat exchangers utilize a heating element to heat the process fluid. [Figures 4.97–4.99](#) are examples of electric heat exchangers.

4.10.2 Heat Recovery Steam Generator

Another type of heat exchanger uses steam circulating in tubes to heat the process fluid. See [Figures 4.100](#) and [4.101](#). A spiral wound exchanger is shown in [Figure 4.102](#). See also [Figure 4.103](#) for a sample spiral heat exchanger.



4.11 HEAT EXCHANGER SELECTION

The proper selection of a heat exchanger involves many factors. It is easy to choose one that will work; however, does it optimize cost of the total system without compromising operating reliability?

4.11.1 Guidelines We Should Follow

- Do not specify or purchase a heat exchanger without considering its effect on the total process.
- Do not make the capital cost of the heat exchanger alone the sole criterion for purchase.
- Acquaint the vendor with details of service and point out that the choice will be made on both initial and operating cost, not just initial capital cost alone.
- Use realistic pressure drop specifications, as this affects size and cost.
- Allow as much pressure loss as economics dictate for the actual system and do not merely reproduce a standard specification that might not apply.
- Remember that the vendor knows the product but knows only as much about the application of the product as the customer can convey.

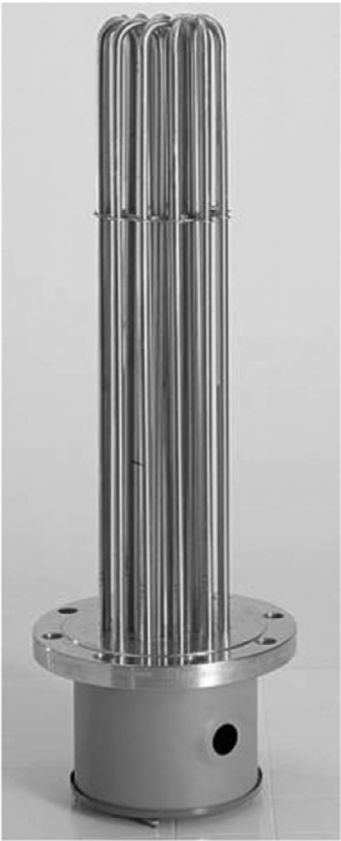
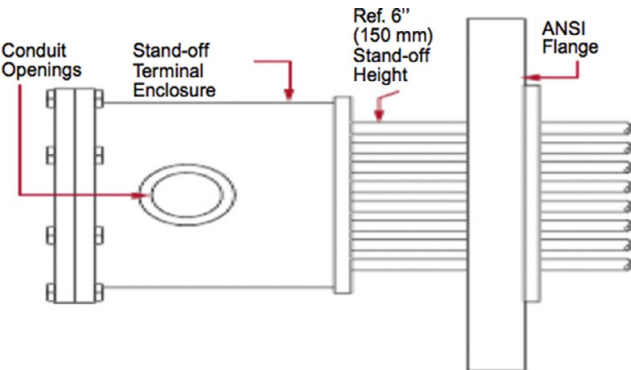


Figure 4.97 Electric heat exchangers.

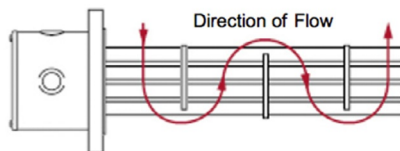
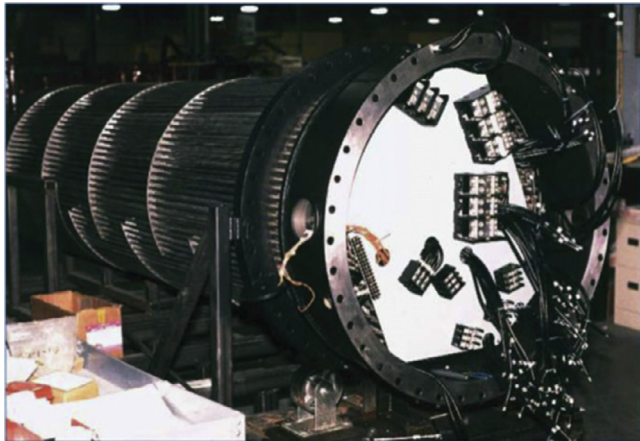


Figure 4.98 Electric heat exchanger.

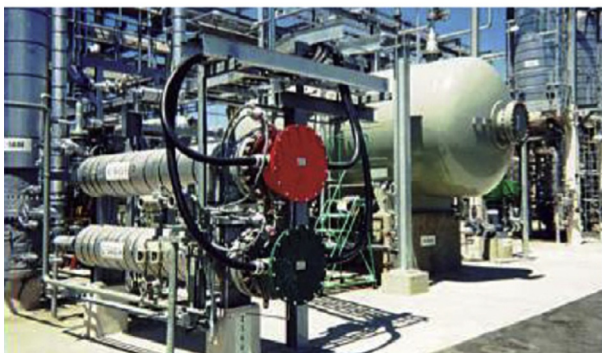
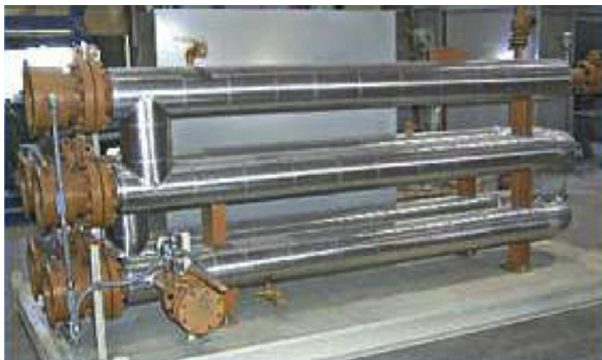


Figure 4.99 Electric heat exchangers.

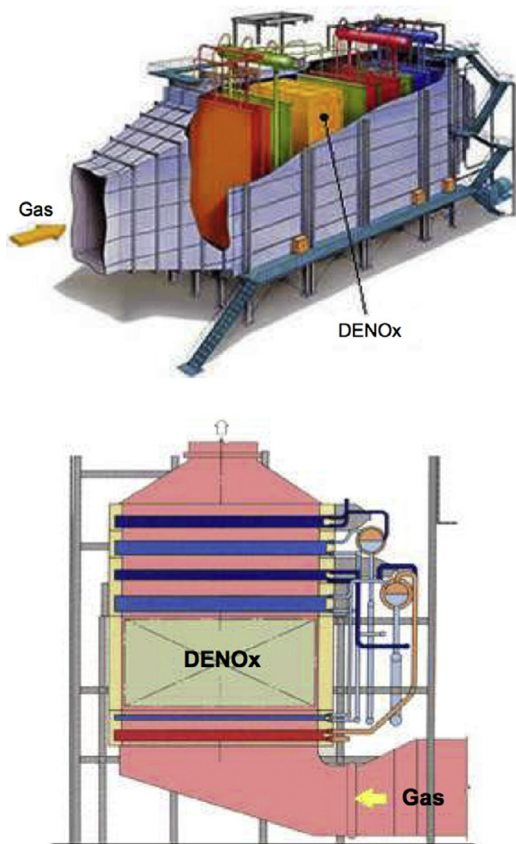


Figure 4.100 Heat recovery steam generator.

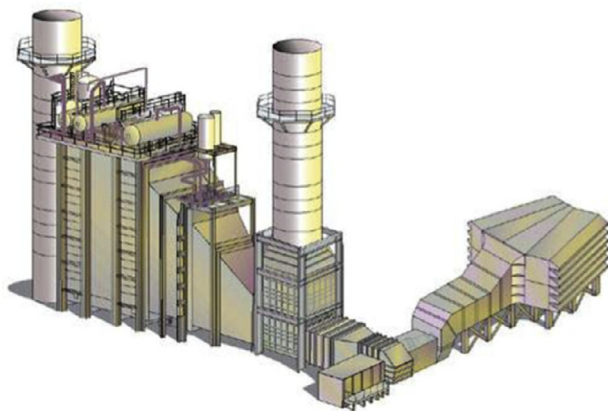


Figure 4.101 Heat recovery steam generator.



Figure 4.102 Spiral wound heat exchanger.

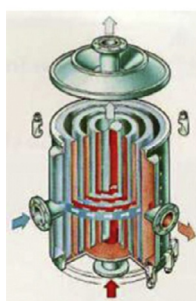
Example 4.2 Line Heater Design

Given

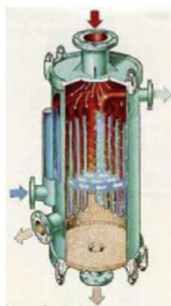
Design a line heater for each of the 10 wells that make up the total 100 MMscfd field rate, that is, each well flows at 10 MMscfd.

Determine

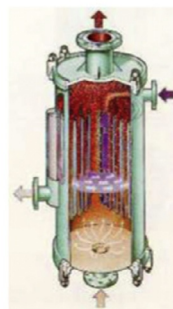
1. Temperature for hydrate formation at 1000 psia
2. Heat duty for a single pass coil downstream of choke



Condenser



Cross flow



Vaporizer

Figure 4.103 Cutaway of a spiral wound heat exchanger.

3. Coil length for a 3-in. Double extra strong (XXS) coil
 - a. Calculate LMTD
 - b. Calculate U
 - c. Choose coil length
4. Fire tube area required and heater size (shell diameter, shell length, fire tube rating, coil length, and number of passes).

Solution

1. Determine the temperature for hydrate formation at 1000 psia
 - a. As shown in [Table 4.9](#), using the vapor equilibrium values from the GPSA Engineering Data Book, or a similar reference, we can determine the hydrate formation temperature at 1000 psia and temperatures of 50 and 70 °F.
Interpolating linearly, S (Y/KV-S). 1.0 at 66.9 °F
 - b. Using the GPSA Engineering Data Book, or equivalent, given specific gravity versus temperature curves at a pressure of 1000 psia, and

Table 4.9 Hydrate Formation Temperature Determination
Y **Values at 1000 psia**

Component	Mole Fraction	50 °F	70 °F
N ₂	0.0144	-	-
CO ₂	0.0403	0.60	-
H ₂ S	0.000019	0.07	0.38
C ₁	0.8555	1.04	1.26
C ₂	0.0574	0.145	1.25
C ₃	0.0179	0.03	0.70
iC ₄	0.0041	0.013	0.21
nC ₄	0.0041	0.145	1.25
iC ₅	0.0020	-	-
nC ₅	0.0013	-	-
C ₆	0.0015	-	-
C ₇₊	0.0015	-	-
	1.0000	$\sum(Y/K_{v-s})=2.226$	$\sum(Y/K_{v-s})=0.773$

specific gravities of 0.6 and 0.7, one can interpolate a hydrate formation temperatures of 60 and 64 °F. At a specific gravity of 0.67, one can interpolate a hydrate formation temperature of 62.8 °F.
From Table 4.9, $S=0.67$

At $S=0.6$, $P=1000$ psia, hydrate temperature = 60°F
At $S=0.7$, $P=1000$ psia, hydrate temperature = 64°F

By interpolation, hydrate temperature at $S=0.67$ is 62.8 °F

2. Determine the process heat duty

The temperature at the outlet of the heater should be between 5 and 15 °F above the hydrate formation temperature.
With a hydrate formation temperature of 62.8 °F, select an outlet temperature of 75 °F.
a. Temperature drop through choke

Flowing tubing pressure, PSIG	4000
Heater inlet pressure, PSIG	<u>1000</u>
DP through choke, PSIG	3000
$\Delta T=79$ °F from Figure 5.7. Curve is based on 20 bbl/MMscf condensate.	
Well has 60 bbl/MMscf, therefore, DT corrected = $79-20=59$ °F	
Accordingly, the heater inlet temperature = $120-59=60$ °F	

b. Gas duty

Flowing pressure, P , psia	1015
P_c , psia 680 Reduced pressure, $P_R = P/P_c$	1.49
Heater inlet temp, $^{\circ}\text{F}$	61
Heater outlet temp, $^{\circ}\text{F}$	75
Average temp, $^{\circ}\text{F}$ $(61 + 75)/2$	68
Average temp, $^{\circ}\text{R}$	528
T_c , $^{\circ}\text{R}$	375
$T_R = T/T_c$	1.41
$q_g = 41.7(\Delta T)C_gQ_g$	

where

q_g	=gas heat duty
Δt	$=t_{\text{out}} - t_{\text{in}}$
Because flow through coil is a constant pressure process, we have:	
$\Delta t = t_{\text{out}} - t_{\text{in}} = 75 - 61$	$=14^{\circ}\text{F}$
Q_g =gas flow, MMscfd	$=10$
C_g	=gas heat capacity, Btu/Mscf- $^{\circ}\text{F}$
Calculation of C_g	
C_g	$=2.64(29)(C) + (\Delta C_p)$
where	
C	=gas-specific heat, Btu/lb $^{\circ}\text{F}$
From Figure 3.32, C at 68°F is $=0.50$	
ΔC_p , from Figure 3.33 (at T_R . 1.41 and $P_R = 1.49$)	$=2.6$
S	$=0.67$
C_g	$=2.64(29)(0.67)(0.50) + (2.6)$
	$=32.51 \text{ Btu/Mscf-}^{\circ}\text{F}$
g_g	$=(41.7)(14)(32.51)(10)$
	$=190 \text{ Mbtu/h}$

c. Oil duty

$$q_o = 14.6(\text{SG})(T_2 - T_1)C_oQ_o$$

where $\text{SG} = 0.77$; Q_o = oil flow rate, BPD, $= (60 \text{ BBL/MMSCF}) (10 \text{ MMscfd})$; $t_2 - t_1 = 75 - 61$, $=14^{\circ}\text{F}$; C_o = oil-specified heat, Btu/lb $^{\circ}\text{F}$. From Figure 3.31 at 68°F (52.3°API), $C_o = 0.48$; $q_o = 14.6(0.77)(14)(0.48)(600)/100$, $=46 \text{ Mbtu/h}$.

d. Water duty $q_w = 14.6 (T_2 - T_1) Q$

Gas is saturated with water at 8000 psig (shut-in BMP) and 224 °F (BHT).

From McKetta Wehe graph, we have

LB water/MMscf of wet gas at reservoir conditions 260

(8000 psig and 224 °F)

LB water/MMscf of wet gas at 1000 psig and 75° 28

Water to be heated 232

$$\begin{aligned}
 \frac{\text{Water quantity}}{\text{quantity}} &= Q_w \\
 &= \left(232 \frac{\text{lb}}{\text{MMscf}} \right) \left(\frac{10 \text{ MMscf}}{62.4 \text{ lb/ft}^3} \right) \left(\frac{7.4 \text{ gal}}{\text{ft}^3} \right) \left(\frac{\text{BBL}}{42 \text{ gal}} \right) \\
 &= 6.6 \text{ BPD} \\
 Q_w &= (14.6)(75 - 61)(6.6) \\
 &= 1.3 \text{ Mbtu/h}
 \end{aligned}$$

e. Total process duty

$$\begin{aligned}
 q &= q_g + q_0 + q_w \\
 &= 90 + 46 + 1.3 \\
 &= 237 \text{ Mbtu/h}
 \end{aligned}$$

3. Calculation of coil length

a. Calculation LMTD

Temperature of bath is 190 °F

$$\Delta t_1 = 190 - 61 = 129$$

$$\Delta t_2 = 190 - 75 = 115$$

$$\begin{aligned}
 \text{LMTD} &= \frac{14}{\log_e \left(\frac{129}{115} \right)} \\
 &= 122^\circ \text{F}
 \end{aligned}$$

b. Calculate U

$$\frac{1}{u} = \frac{1}{h_o} + R_0 + \frac{L}{K} + R_i + \frac{A_0}{h_{iA_1}}$$

Uses $R_o + R_i = 0.003 \text{ h ft}^2 \text{ }^\circ\text{F/Btu}$

$$h_0 + 116 \left[\frac{k^3 C^2 \rho \beta \Delta t}{\mu d_0} \right]^{0.25}$$

where $K=0.39$ (Figure 4.26); $C_p=1$ Btu=lb °F (Figures 4.24, 4.31, and 4.33); $\rho=60.35$ lb/ft³ (Table, 1/0.01657); $\beta=0.0024$ 1/°F (Table 4.5); $\mu=0.32$ cp; $\Delta t=122$ °F; $d_o=3.5$ in.

$$h_0 = 116 \left[\frac{(0.39)^3 (1)^2 (60.35) (0.0024) (122)}{(0.32)(3.5)} \right]^{0.25}$$

$$= 114 \text{ Btu/h ft}^2 \text{ P}$$

For 3-in. XXH pipe A-106B

L	=0.60 in.-0.50 ft (Table)
K	=0 Btu/h ft °F (Table)
	=(for 90 CU and 10 NI)
A_0	=0.916 ft ² /ft (Table)
A_i	=0.602 ft ² /ft (Table)

$$h_i = \frac{0.022K}{D} \left(\frac{DG}{\mu_e^{0.8}} \right) \left(\frac{C_{\mu e}}{K^{0.4}} \right) \left(\frac{\mu_e}{\mu_{ew}} \right)^{0.16}$$

where $D=2.30$ in = 1.92 ft (Table); $A=\pi D^2/4=0.0289$ ft²;
 $K=0.017$ Btu/h ft °F 1.25 (Figure), = 0.021 Btu/h ft °F.

$$\text{Gas Flow} = (10 \text{ MMscf}) \left(\frac{D}{24 \text{ h}} \right) \left(\frac{\text{lb/mol}}{379 \text{ scf}} \right) \left(\frac{19.4 \text{ lb}}{\text{lb/mol}} \right)$$

$$= 21,238 \text{ lb/h}$$

$$\text{Oil Flow} = \left(\frac{608 \text{ BBL}}{D} \right) \left(\frac{D}{24 \text{ h}} \right) \left(\frac{350 \text{ lb}}{\text{BBL}} \right) (0.77)$$

$$= 6827 \text{ lb/h}$$

$$\text{Water Flow} = \left(\frac{6.6 \text{ BBL}}{D} \right) \left(\frac{D}{24 \text{ h}} \right) \left(\frac{350 \text{ lb}}{\text{BBL}} \right)$$

$$= 96 \text{ lb/h}$$

G = mass velocity of fluid

$$= \frac{(21,328) + (6827) + (96)}{0.0299}$$

$$= 977,560 \text{ lb/h - ft}^2$$

$$\begin{aligned}
 c &= \left(\frac{32.51 \text{ btu}}{\text{Mscf}^\circ\text{F}} \right) \left(\frac{6827 \text{ lb}}{28,252 \text{ lb}} \right) + \left(\frac{1 \text{ BTU}}{\text{lb}^\circ\text{F}} \right) \left(\frac{96 \text{ lb}}{28,252 \text{ lb}} \right) \\
 &+ \left(\frac{0.48 \text{ btu}}{\text{lb}^\circ\text{F}} \right) \left(\frac{6827 \text{ lb}}{28,252 \text{ lb}} \right) + \left(\frac{1 \text{ btu}}{\text{lb}^\circ\text{F}} \right) \left(\frac{96 \text{ lb}}{28,252 \text{ lb}} \right) \\
 &= 0.60 \text{ Btu/lb}^\circ\text{F}
 \end{aligned}$$

$$\mu = 0.0134 \text{ cp (at } 60^\circ\text{F)} (2.4) = 0.0322 \text{ lb/h ft}$$

$$\mu_w = 0.0142 \text{ cp (at } 129^\circ\text{F)} (2.4) = 0.0341 \text{ lb/h ft}$$

$$\begin{aligned}
 0h_i &= \frac{(0.022)(0.021)}{0.192} \left(\frac{(0.192)(977.560)}{0.0322} \right)^{0.8} \\
 &+ \left(\frac{(0.6)(0.0322)}{0.021} \right)^{0.4} \left(\frac{0.0322}{0.034} \right)^{0.16} \\
 &= 595 \text{ Btu/h ft}^2^\circ\text{F}
 \end{aligned}$$

$$\frac{1}{u} = \frac{1}{114} + 0.003 + \left(\frac{0.05}{30} \right) + \left(\frac{0.9162}{(595)(0.6021)} \right)$$

U	$= 62.5 \text{ Btu/h ft}^2^\circ\text{F}$
Estimated U from Figure 4.92	
U	$= 106 \text{ Btu/h ft}^2^\circ\text{F}$
Use U	$= 62.5 \text{ Btu/h ft}^2^\circ\text{F}$

c. Calculate coil length

$$L = \frac{12q}{\pi(\text{LMTD})Ud}$$

where $q = 237 \text{ MBtu/h}$ (total process heat duty); $L = 122^\circ\text{F}$;
 $U = 62.5 \text{ Btu/h ft}^2^\circ\text{F}$; $d = 3.5 \text{ in}$:

$$\begin{aligned}
 L &= \frac{(12)(237,000)}{\pi(122)(62.5)(3.5)} \\
 &= 39.9 \text{ ft}
 \end{aligned}$$

4. Calculate fire-tube area required

For heat transfer to water use $10,000 \text{ Btu/h ft}^2$

Flux rate

$$\begin{aligned}
 A &= \frac{\frac{237,000 \text{ btu}}{\text{h}}}{\frac{10,000 \text{ btu}}{\text{h}} - \text{ft}^2} \\
 &= 23.7 \text{ ft}^2
 \end{aligned}$$

Estimate shell size

Assuming a 10'-0" shell, then four passes of 3" XXH are required. This will require a 30-in. OD shell for the coils and fire tube.

5. Summary of line heater size

Heater duty	250 MBtu/h
Coil size	300XXH
Minimum coil length	33.9 ft
Minimum fire-tube area	23.7 ft
Shell size 3	0" OD \times 10'-0' F/F

Example 4.3 Shell-and-Tube Heat Exchanger Design

Given

Inlet	100 MMscfd at 0.67 SG (Table) $T_1 = 175^\circ\text{F}$ $P_1 = 1000$ psig Water vapor in gas = 60 lb/MMscf 15 bbl water/MMscf
Outlet	$T_2 = 100^\circ\text{F}$ $P_2 = 990$ psig Water vapor in gas = 28 lb/MMscf
Seawater	$T_3 = 75^\circ\text{F}$ Limit temperature rise to 10°F

Use 1-in. OD 10 BWG tubes on 1¼-in. pitch

Design a seawater cooler to cool the total stream from the example field in its later stages of life from a flowing temperature of 175°F to a temperature of 100°F to allow further treating.

Determine

1. Calculate water vapor condensate
2. Calculate heat duty
3. Determine seawater circulation rate
4. Pick a type of exchanger and number of tubes
5. Check to ensure water velocity in tubes does not exceed 15 ft/s

Procedure

Calculate heat duty

Determine fluid in shell/tube

Assume/calculate overall heat transfer coefficient

Choose

- (a) number of shell passes
- (b) number of tube passes
 - Correct LMTD
 - Select tube diameter and length
 - Calculate the number of tubes
 - Determine shell diameter
 - Determine type of exchanger
 - Determine tube velocity

Solution

1. Calculate free water and water vapor inlet flow rates

$$\begin{aligned}\text{Free water} &= (100 \text{ MMscfd})(15 \text{ bbl/MMscfd}) \\ &= 1500 \text{ bwpd}\end{aligned}$$

$$\begin{aligned}\text{Water vapor condensed} &= \left(\frac{(60 - 28) \text{ lb}}{\text{MMscf}} \right) \left(\frac{100 \text{ MMscf}}{d} \right) \\ &= 3200 \text{ lb/day}\end{aligned}$$

$$\begin{aligned}Q_w &= (3200 \text{ lb/d}) \left(\frac{\text{bbl}}{350 \text{ lb}} \right) \\ &= 9 \text{ bwpd}\end{aligned}$$

Water flow rate outlet

9	bwpd
<u>1500</u>	<u>bwpd</u>
1509	bwpd

2. Calculate heat duty

- a. Gas duty

T_1	$= 635^\circ \text{R} (175^\circ \text{F inlet})$
T_2	$= 560^\circ \text{R} (100^\circ \text{F outlet})$
T_{AV}	$= 597.5^\circ \text{R}$
P_C	$= 680 \text{ psia}$
P_R	$= P/P_C = 1.47 (1000/680)$
T_C	$= 370^\circ \text{R}$
T_R	$= T_{\text{AVG}}/T_C = 1.62 (597.5/370)$
q_g	$= 41.7 \Delta T C_g Q_g \text{ (sensible heat of natural gas)}$
C_g	$= 2.64 (29 \text{ SG} + \Delta \text{CP})$
C	$= 0.528 \text{ Btu/lb } ^\circ \text{F}$

$$\begin{aligned}
 \Delta C_p &= 1.6 \text{ Btu/lb-mol } ^\circ\text{F} (P_R \text{ \& } T_R) \\
 S &= 0.67 \text{ (Table)} \\
 C_g &= 2.64 [(29)(0.67)(0.528) + 1.6] \\
 &= 31.3 \\
 q_g &= 41.7(100 - 175)(31.3)(100) \\
 &= 9,789,000 \text{ Btu/h}
 \end{aligned}$$

b. Condensate duty

$$\begin{aligned}
 q_o &= 14.6 \text{ (S.G.) } \Delta T C_o Q_o \\
 C_o &= 0.535 \text{ Btu/lb } ^\circ\text{F (Figure)} \\
 q_o &= 14.6 (0.77) (100 - 175)(0.535)(6000) \\
 &= -2,707,000 \text{ Btu/h}
 \end{aligned}$$

c. Free water duty

$$\begin{aligned}
 q_w &= 14.6 \Delta T Q_w \\
 &= (14.6)(100 - 175)(1509) \\
 &= 1,652,000 \text{ Btu/h}
 \end{aligned}$$

d. Water latent heat duty

$$\begin{aligned}
 q_{lh} &= (3200 \text{ lb/D})(1/24) \\
 &= 133 \text{ lb/h} \\
 \lambda &= -996.3 \text{ Btu/lb (Table, } 170 ^\circ\text{F)} \\
 q_{lh} &= (133) (-996.3) \\
 &= -133,000 \text{ Btu/h}
 \end{aligned}$$

e. Total heat duty

$$\begin{aligned}
 q &= q_g + q_o + q_w + q_{lh} \\
 &= -9,789,000 - 2,707,000 - 1,652,000 - 133,000 \\
 &= -14,281,000 \text{ Btu/h}
 \end{aligned}$$

3. Water circulation rate

$$q_w = 14.6(t_2 + t_1)Q_w$$

Rearranging and solving for Q_w

$$Q_w = \frac{q_w}{14.6(t_2 - t_1)}$$

Limit DT for water to 10 °F to limit scale

$$\begin{aligned} Q_w &= (14.3 \times 10^6)(10) \\ &= 97,945 \text{ bwpd} \\ &= 2858 \text{ gpm} \end{aligned}$$

4. Heat exchanger type and number of tubes

Choose TEMA R because of large size (offshore)

Select type AFL because of low temperature change and LMTD correction factor

A	channel and removal cover (used commonly with fixed tube sheet)
F	two-pass shell with longitudinal baffle
L	fixed tube sheet
	least expensive
	fewer gaskets
	individual tubes can be replaced
	used in clean shell fluids and low differential temperatures (<200 °F)
	shell cannot be cleaned and inspected
	bundle cannot be replaced

If brackish water, use higher nickel content for tube material, therefore, 70/30 C_UN_i. Because the water is corrosive and may deposit solids, flow water through tubes.

$U = 90 \text{ Btu/h } ^\circ\text{F}$ (Table) (water with 1000 PSI gas)

Calculate LMTD

t_1	=175
t_2	=100
t_3	=75
t_4	=85
Δt_1	=175 – 85 = 90
Δt_2	=100 – 75 = 25

$$\begin{aligned} \text{LMTD} &= \left[\frac{90 - 25}{\log_e \left(\frac{90}{25} \right)} \right] \\ &= 50.7^\circ\text{F} \end{aligned}$$

Correction factor (from Figure)

$$P = \frac{85 - 75}{175 - 75}$$

$$= 0.1$$

F

MTD

$$0.95 (50.7)(0.95) \ 48.2^\circ\text{F}$$

$$R = \frac{175 - 100}{85 - 75}$$

$$= 7.5$$

$$= 0.95$$

$$= (50.7)(0.95)$$

$$= 48.2^\circ\text{F}$$

Calculate number of tubes

$$N = \frac{q}{UA'(LMTD)L}$$

Assume

L	= 40 ft
A	= 0.2618 ft ² /ft (Table)
	= 315 tubes

From Table for 1-in. OD, 1¼ in. SQ pitch, fixed tube sheet, two passes, shell ID = 29 in.

If, Assume L = 20 ft (instead of 40 ft)

Then N = 629 tubes

Shell ID = 39 inches

Use 39-in. ID × 20 ft long with 682-in. OD, 10 BWG tubes 1¼-in., square pitch with two tube passes.

Size 39-240 type AFL

5. Check water velocity

$$V = 0.012 \frac{Q_w}{d^2}$$

There are four passes. Thus, 682/2 tubes are used in each pass.

$$Q_w \text{ per tube} = \frac{(4)(97,945)}{682}$$

$$= 227 \text{ bwpd}$$

$$D = 0.732 \text{ in. (from table)}$$

$$V = \frac{(0.012)(574)}{(0.732)^2}$$

$$= 6.4 \text{ ft/s, which is } < 15 \text{ ft/s}$$



4.12 COMMENTS ON EXAMPLE 4.3

Once the heat exchanger has been chosen and the flow per tube, a more precise “U” can be determined. More than 30% of the heat duty was required to cool the water and condensate. If liquids had first been separated:

- Less heat duty is required
- Smaller heat exchanger is needed
- Lower seawater flow rate

Therefore, a cooler is usually placed downstream of the first separator. Often, an aerial cooler is used in this service. Outlet temperature of 100 °F may be sufficient if only compression and dehydration are required.

If gas treating is required (H₂S or CO₂) the amine unit will heat gas 10–20 °F. Thus, gas dehydration would be difficult at elevated temperature. Therefore, it is better to cool gas lower initially so that it is still below 110 °F at glycol dehydration. Often, this is not possible because cooling water is not available or ambient air conditions are in the 95–100 °F range. This type of service may require an aerial cooler to cool the gas before treating it and before dehydration.



EXERCISES

1. Match the following heat transfer equations with their description.

a	Convection ____	$q = UA\Delta T$
b	Conduction ____	$q = hA\Delta T$
c	Radiation ____	$q = \Delta T / \Sigma R$
d	Overall heat transfer equation ____	$q = A\sigma T^4$
e	Thermal resistance ____	$q = (kA/L) \Delta T$

2. What effect do fouling factors (r_0 and r_i) have on the overall heat transfer coefficient?
3. What is the advantage of high velocity in heat exchangers?
4. What are the three assumptions used in the derivation or LMTD?
 - a. _____
 - b. _____
 - c. _____
5. Double-pipe exchangers are best suited for applications involving large heat transfer surfaces. True or False

6. What is the main difference between floating head and fixed tube sheet exchangers?_____
7. What is the different between tube pitch and tube clearance?
8. What are the purposes of baffles in shell-and-tube exchangers?_____
9. TEMA Class “R” heat exchangers are more heavy duty than Class “C” heat exchangers. True or False
10. What is the purpose of fins in air-cooled exchangers?
11. Compare the fuel efficiency of a fired heater using only a radiant section with that of a fired heater using a radiant and convection section.
12. Indicate which type of heat transfer equipment is described: double pipe, shell-and-tube, plate-fin, plate-and-frame, air cooler, fired heater, or waste heater
 _____Finned tube-and-fan system
 _____Pipe within a pipe
 _____Radiant section
 _____Most are equipped with baffles
 _____Bundle of tubes within a shell
 _____Clean services only
 _____Convection section only
 _____Shield tubes
 _____True counter-current
 _____Very close approaches
 _____Efficiency is dependent on fan performance
 _____Uses heater or turbine exhaust
13. **Given:**

Inlet condition		
T_1		=165 °F
P_1		=600 psig
Moisture content		=70 lb H ₂ O/MMSCF
Q_w		=20 BBL/MMSCF
Outlet conditions		
T_2		=100 °F
P_2		=585 psig (minimum)
Moisture content		=25 lb H ₂ O/MMSCF
Seawater		
T_3		=72 °F

Determine:

Design a shell-and-tube heat exchanger that will cool 175 MMSCFD ($S=0.65$) of gas from a flowing temperature of 165° to 100°F .

Assume 1-in. OD 10 BWG tubes on $1\frac{1}{4}$ -in. Δ pitch are available and the process is located offshore.

14. What type of heat exchanger is the workhorse for separation, heating, and cooling?
15. What is the difference, if any, between co-current (parallel-flow) heat exchangers and counter-current flow heat exchangers?
16. Heat exchangers are not susceptible to problems because they have no moving parts. True or False
17. What has the Tubular Exchanger Manufacturers Association developed for describing shell-and-tube heat exchangers?
18. For shell-and-tube heat exchangers, what is the typical shell type choice? Why?
19. Explain how a cooling tower, which is a specialized heat exchanger, transfers heat.
20. Plate-and-frame exchangers are used mostly in upstream installations. True or False
21. What TEMA size would a 40-ft straight length U-tube bundle, 6-ft shell diameter, with a single shell pass and a removable shell cover be?
22. What is the most important distinguishing feature describing heat exchangers?
23. During exchanger design, what is the setting of the tube count per pass based on?
24. Why is it difficult to clean heavy deposits from the inside on U tubes?
25. Compact heat exchangers get their name from their small size. True or False?



Hydrate Prediction and Prevention



5.1 OBJECTIVES

To prepare natural gas for sale, its undesirable components (water, H_2S , and CO_2) must be removed. Most natural gas contains substantial amounts of water vapor due to the presence of connate water in the reservoir rock. At reservoir pressure and temperature, gas is saturated with water vapor. Removal of this water is necessary for sales specifications or cryogenic gas processing. Liquid water can form hydrates, ice-like solids that can plug flow or decrease throughput. Primary concerns in surface facilities are determining the water content of the gas and conditions under which hydrates will form.

Predicting the operating temperatures and pressures at which hydrates form and methods of hydrate prevention are discussed in this section.



5.2 OVERVIEW

5.2.1 Dew Point

The dew point is the temperature and pressure at which the first drop of water vapor condenses into a liquid. It is used as a means of measuring the water vapor content of natural gas. As water vapor is removed from the gas stream, the dew point decreases. Keeping the gas stream above the dew point prevents hydrates from forming and prevents corrosion from occurring.

5.2.2 Dew-Point Depression

Dew-point depression is the difference between the original dew point and the dew point achieved after some of the water vapor is removed. It is used to describe the amount of water that needs to be removed from the natural gas to establish a specific water vapor content.

5.2.3 Why Dehydrate?

Dehydration refers to removing water vapor from a gas to lower the stream's dew point.

If water vapor is allowed to remain in the natural gas, it will reduce the efficiency and capacity of a pipeline, cause corrosion that will eat holes in the pipe or vessels through which the gas passes, and form hydrates or ice blocks in pipes, valves, or vessels. Dehydration is required to meet gas sales contracts (dependent on ambient temperatures). Some examples include:

- Southern United States, Southeast Asia, southern Europe
- West Africa, Australia 7 lb/MMSCFD
- Northern United States, Canada, northern Europe, northern and central Asia 2–4 lb/MMSCFD
- Cryogenic (turbo expander plants) 0.05 lb/MMSCFD

Solid bed adsorption units are used where very low dew points are required.



5.3 WATER CONTENT OF GAS

5.3.1 Introduction

Liquid water is removed by gas-liquid and liquid-liquid separation. The capacity of a gas stream to hold water vapor is a function of the gas composition, affected by the pressure and temperature of the gas, and is reduced as the gas stream is compressed or cooled.

When a gas has absorbed the limit of its water holding capacity for a specific pressure and temperature, it is said to be saturated or at its dew point. Any additional water added at the saturation point will not vaporize but will fall out as free liquid. If the pressure is increased and/or the temperature is decreased, the capacity of the gas to hold water will decrease, and some of the water vapor will condense and drop out.

Methods of determining the water content of gas include:

- Partial pressure and partial fugacity relationships
- Empirical plots of water content versus P and T

Corrections to the empirical plots are presented to adjust the water content for the presence of contaminants such as hydrogen sulfide, carbon dioxide, and nitrogen.

5.3.2 Partial Pressure and Fugacity

Applying Raoult's law of partial pressures to water, we have

$$\gamma_w = P_v x \quad (5.1)$$

where y_w = mol fraction of water in the vapor phase; P_v = vapor pressure of water at system temperature; x_w = mol fraction of water in the liquid water phase, =1.0.

Liquid mol fraction can be taken as unity because of the immiscibility of the liquid phases.

Therefore, for a known pressure and water vapor pressure, the mol fraction water in the vapor phase can be determined from Equation (5.1). Application of Equation (5.1) is

- Valid only at low pressure where the ideal gas law is valid and
- Recommended for system pressures up to 60 psia (4 barg)

5.3.3 Empirical Plots

Empirical plots are based on lean, sweet natural gas. The log of water content (w) is plotted versus P and T . Plots approximate a straight line at a given pressure. The water content shown is the maximum that gas can hold at the P and T shown. It is fully saturated, that is, relative humidity is 100%. The temperature is the water dew point temperature of the gas at the concentration and pressure shown.

Numerous correlations are available to determine the water content of a natural gas stream. The McKetta and Wehe correlation provides satisfactory results for most applications when used to determine the water content of a sweet natural gas stream that contains over 70% methane (Figure 5.1). The accuracy is $\pm 5\%$ (probably more accurate than the data toward which the correlation is being applied).

As H_2S and CO_2 content increases, the accuracy decreases. It is good practice to make corrections for these contaminants even though it may be small when concentrations and pressure are low.

Comparison of the water content at various points in a system allows one to determine the water loading for dehydration and how much water has been condensed as liquid in the pipeline, which is available to form hydrates and is the root cause of many corrosion-erosion problems.

5.3.4 Sour Gas Correlations

5.3.4.1 Weighted-Average Method

Correlation employs a weighted average to determine the water content of a gas stream containing sour gas. In this correlation, the water content of the pure sour component is multiplied by its mol fraction in the mixture. The following equation can be used:

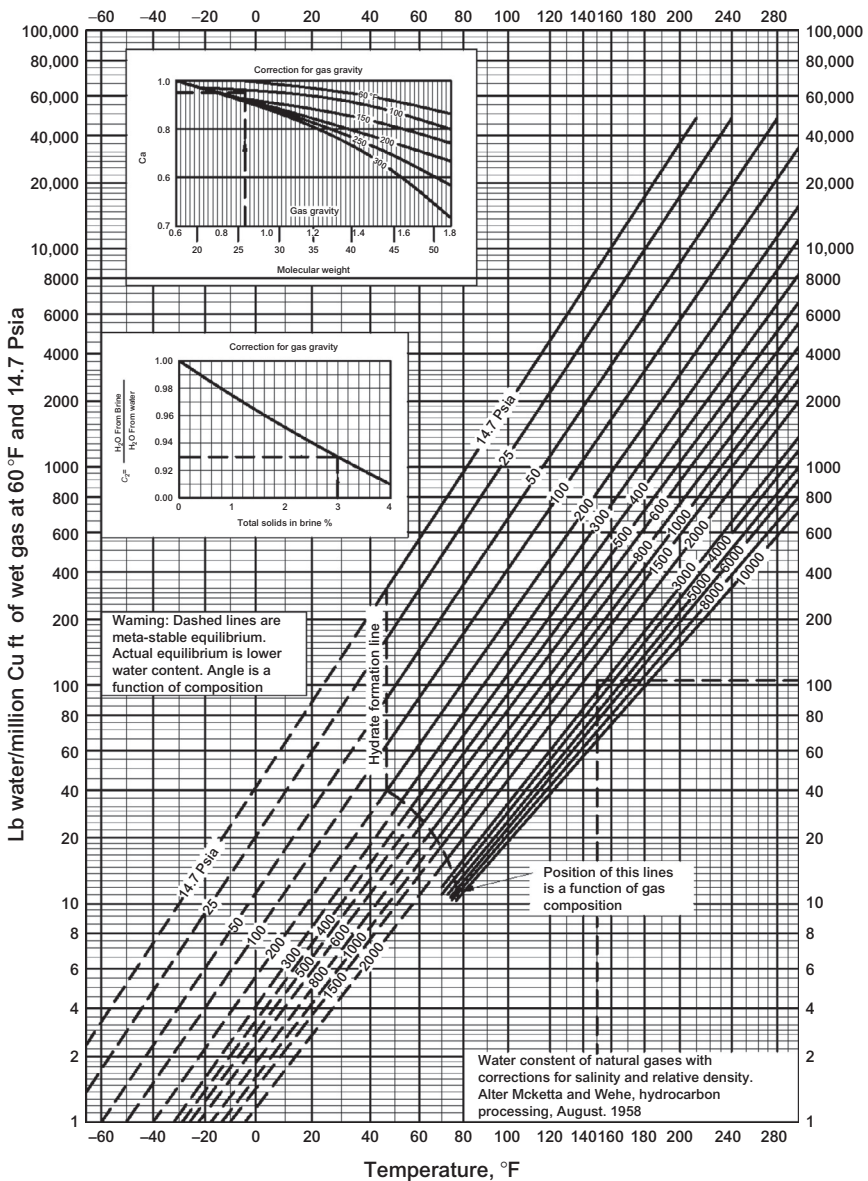


Figure 5.1 Water content of sweet, lean natural gas—McKetta-Wehe.

$$W = \gamma W_{hc} + \gamma_1 W_1 + \gamma_2 W_2 \tag{5.2}$$

where W =water content of gas; W_{hc} =water content of hydrocarbon part obtained from McKetta-Wehe plot; W_1 =water content of CO_2 obtained from appropriate empirical plot; W_2 =water content of the H_2S obtained

from appropriate empirical plot; $\gamma = 1 - (\gamma_1\gamma_2)$; γ_1 = mol fraction of CO_2 ; γ_2 = mol fraction of H_2S .

Figures 5.2 and 5.3 are correlations that estimate the effective water content of a pure sour component. These correlations should be used when the gas stream contains a combined concentration of more than H_2S and CO_2 combined.

5.3.4.2 Sharma Correlation

The Sharma correlation utilizes Equation (5.2) and is based on the data obtained by Sharma.

Figures 5.4 and 5.5 were obtained by cross-plotting and smoothing Sharma's binary data for methane, CO_2 , and H_2S .

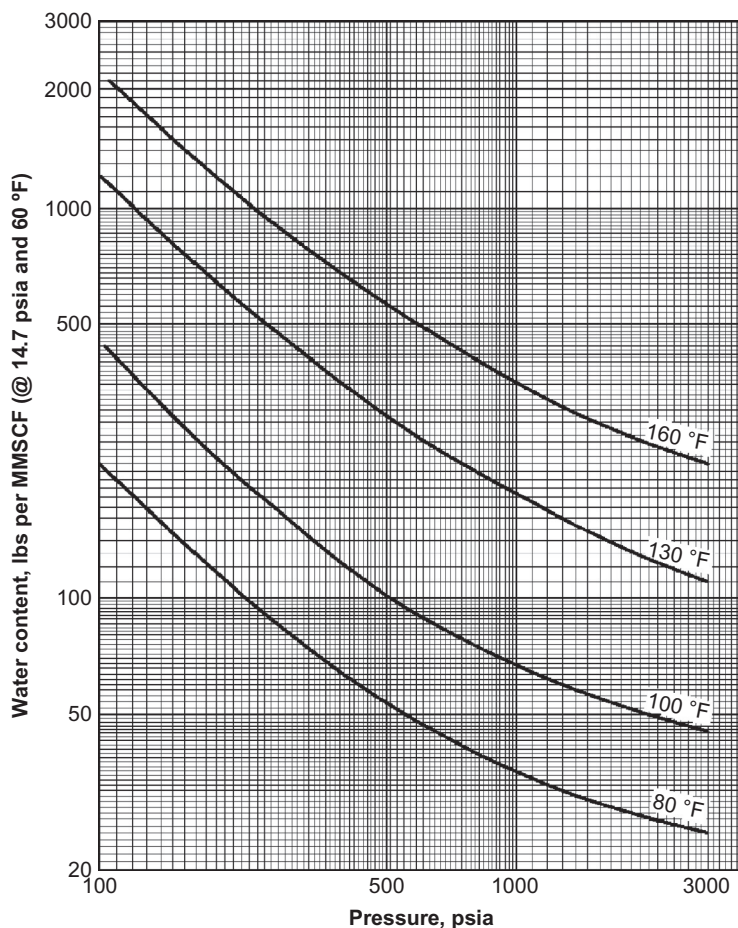


Figure 5.2 Effective water content of CO_2 in saturated natural gas mixtures.

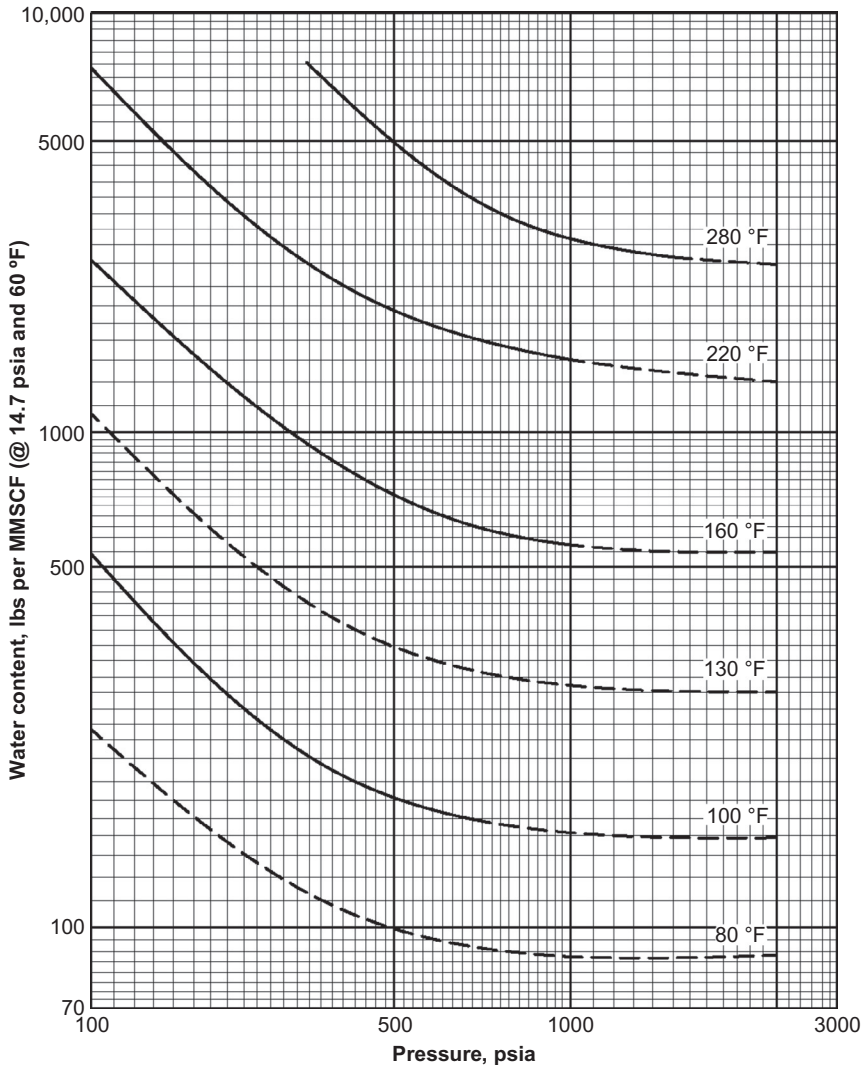


Figure 5.3 Effective water content of H_2S in saturated natural gas mixtures.

5.3.4.3 SRK Sour Gas Correlation

The charts in Figure 5.6 were calculated from the SRK equation of state with the following assumptions:

- The hydrocarbon portion of the gas was methane.
- CO_2 had 75% of the water content of H_2S at the same conditions. One must multiply the percent CO_2 by 0.75 and add the result to the percent H_2S .

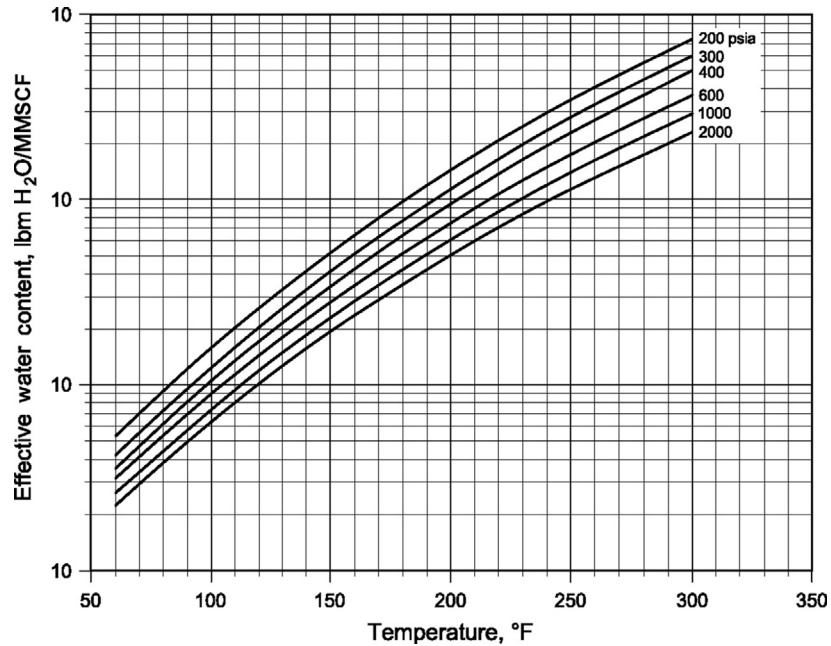


Figure 5.4 Water content of CO₂—Sharma.

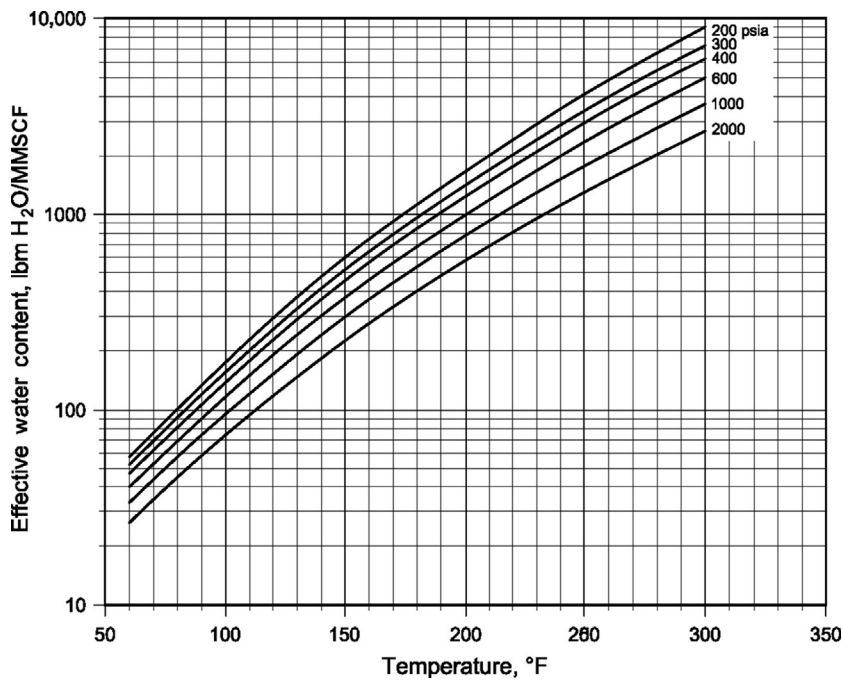


Figure 5.5 Water content of H₂S—Sharma.

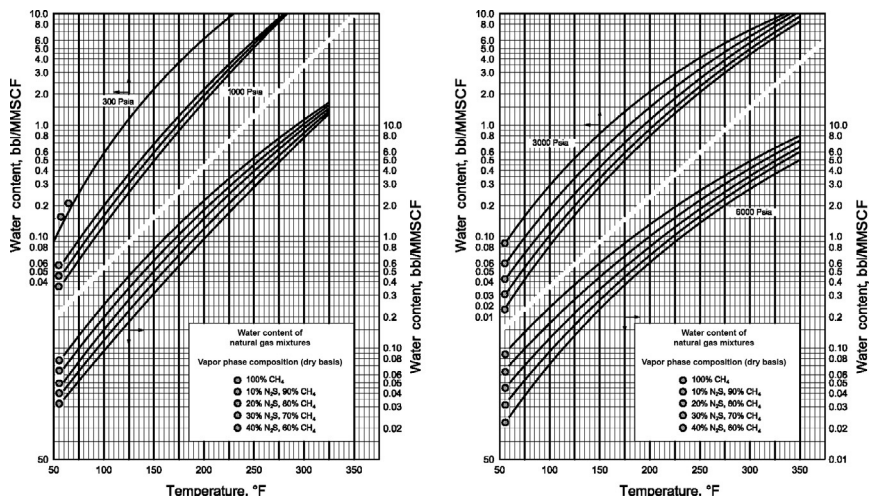


Figure 5.6 Sour gas correlation—SRK.

The water content shown in API bbl/MMSCF can be converted as follows:

lbm/MMSCF: $(350)(\text{bbl/MMSCF})$. The correlation is a “quick look” way to estimate sour gas content.

5.3.5 Effect of Nitrogen and Heavy Ends

Nitrogen holds less water than methane. Pressures up to 1000 psia (69 bara) water content of nitrogen is 5–10% less than methane. Deviation increases as pressure increases. Including nitrogen as a hydrocarbon is practical and offers somewhat of a safety factor. Presence of heavy ends tends to increase the water capacity of the gas. Deviation is relatively small at normal system pressures. The effects of nitrogen and heavy ends tend to cancel each other out in most systems.

Example 5.1: Calculation of Water Content in a Sour Gas Stream Determine

Calculate the saturated water content of a gas with the analysis shown below at 1100 psia (759 bara) and 120 °F (49 °C).

Composition	y_i
N ₂	0.0046
CO ₂	0.0030
H ₂ S	0.1438

C_1	0.8414
C_2	0.0059
C_3	0.0008
iC_4	0.0003
nC_4	0.0002
	1.0000

Solution

1. From McKetta-Wehe plot read
 - a. $W = 97$ lbm/MMSCF
2. From “effective water” content plots for CO_2 and H_2S , read
 - b. $W_1 = 130$, $W_2 = 230$
 Substitute values into Equation (5.2):

$$\begin{aligned}
 W &= (0.8532)(97) + (0.003) + (130) + (0.1438)(230) \\
 &= 116 \text{ lbm/MMSCF}
 \end{aligned}$$

3. From Sharma plots, $W_1 = 120$, $W_2 = 150$
 Substitute in Equation (5.2):

$$\begin{aligned}
 W &= (0.8532)(97) + (0.003)(120) + (0.1438)(150) \\
 &= 105 \text{ lbm/MMSCF}
 \end{aligned}$$

4. Effective percent of

$$\begin{aligned}
 \text{H}_2\text{S} &= (\% \text{CO}_2)(0.75) + (\% \text{H}_2\text{S}) \\
 &= (0.3)(0.75) + (14.38) \\
 &= 14.6\%
 \end{aligned}$$

5. From SRK correlation plot, we must convert bbl/MMSCF to lbm/MMSCF

$$\begin{aligned}
 W &= (350)(\text{bbl/MMSCF}) \\
 &= (350)(0.31) \\
 &= 109 \text{ lbm/MMSCF}
 \end{aligned}$$

Note that the water content is greater using Equation (5.2) than from McKetta-Wehe plot.

A value of 116 is not likely but could happen. Do not treat any one number as sacred. Look at the range when determining water content.

5.3.6 Applications

Correlations are used to:

- Determine the water content for dehydration calculations
- Determine how much water, if any, will condense from the gas—this involves considering disposal, corrosion-erosion, and hydrate inhibition

5.3.7 Amount of Water Condensed

The designer needs to make certain estimates are on the safe side of the possible range of values.

The additional capital expenditure will almost always be trivial. There is a tendency to predict flowing temperatures lower than they typically turn out to be. The reason for this is the quality of the data used. Much of the data is obtained from a drill stem test, which is mediocre at best. Well flowing temperature usually stabilizes upward after a few months in service.

The McKetta-Wehe plot is based on the log scale and thus a small change in temperature will result in a larger change in water content. For example, a 10% change in temperature results in a 33% increase in water content.

A common cause of poor dehydrator performance is under-predicting the water load.



5.4 GAS HYDRATES

5.4.1 What Are Gas Hydrates?

Gas hydrates are complex lattice structures composed of water molecules in a crystalline structure. Hydrates resemble dirty ice but have voids into which gas molecules will fit. The most common compounds found in the lattice voids are water, methane, and propane or water, methane, and ethane.

The physical appearance resembles a wet, slushy snow until they are trapped in a restriction and exposed to differential pressure, at which time they become very solid structures, similar to compacting snow into a snowball.

5.4.2 Why Is Hydrate Control Necessary?

Gas hydrates accumulate at restrictions in flow lines, chokes, valves, and instrumentation and accumulate into the liquid collection section of vessels. The hydrates plug and reduce line capacity, cause physical damage to chokes and instrumentation, and cause separation problems.

5.4.3 What Conditions Are Necessary to Promote Hydrate Formation?

Correct pressure and temperature and “free water” should be present, so that the gas is at or below its water dew point. If “free water” is not present, hydrates cannot form.

5.4.4 How Do We Prevent or Control Hydrates?

Hydrates can be prevented or controlled by the following actions:

- Adding heat to the gas to keep the temperature above the hydrate formation temperature
- Lowering the hydrate formation temperature with chemical inhibition
- Dehydrating the gas so that water vapor will not condense into “free water”
- Designing the process to melt hydrates by using a LTX unit



5.5 PREDICTION OF OPERATING TEMPERATURE AND PRESSURE

5.5.1 Wellhead Conditions

Temperature and pressure of a gas stream at the wellhead are important factors in determining whether hydrates will form when gas is expanded into the flow lines.

Temperatures at the wellhead increase as the flow rate increases and the pressure decreases.

Thus, wells that initially flowed under conditions causing hydrates to form in downstream equipment may decline out of the hydrate formation region as the reservoir depletes and the wellhead pressure drops.

Hydrate formation can sometimes be prevented if the flow rate from a well is maintained above some minimum rate. This is an effective use of reservoir energy that would otherwise be lost in the pressure drop across a choke.

5.5.2 Flowline Conditions

The cooling of a gas in a flowline due to heat loss to the surroundings (ground, water, or air) can cause the gas temperature to drop below the hydrate formation temperature. Records of flowline temperatures and pressures are needed to determine the best locations to affect pressure drops or install heaters.

5.5.3 Calculation of Temperature and Pressure at the Wellhead

Numerous computer programs are available that calculate the temperature and pressure of a gas stream at the wellhead and predict changes that will occur as the reservoir depletes. Calculation by hand is tedious and requires numerous iterations.

5.5.4 Calculation of Flowline Downstream Temperature

The conduction-convection equation can be used to calculate the downstream temperature of a flowline (T_d)

$$T_d = T_g + \frac{T_u - T_g}{e^x} \quad (5.3)$$

where T_d = flowline downstream temperature, °F.

$$x = 24 \frac{(\pi D U L)}{(Q C_p)}$$

where D = flowline OD, ft.; U = heat transfer coefficient, Btu/h/ft.²/°F, =Table 5.1; L = flowline length, ft.; Q = gas flow rate, MSCFD; C_p = specific heat factor, Btu/MCF/°F = 26,800 normally used (values in Table 5.2, multiplied by 1000 may yield more accurate results); $e = 2.718$; T_u = upstream gas temperature, °F (It could be the wellhead temperature (T_{WH}) if no choke or heater is used, or it could be the temperature downstream of a heater.); T_g = ground temperature, °F, =Table 5.3.

Table 5.1 Heat Transfer Coefficients (U) for Various Bare Pipe Conditions

Type Cover	Cover Condition	Depth of Cover (in.)	(Btu/h/ft. ³ /°F)
	Dry	24	0.25-0.40
	Moist	24	0.50-0.60
	Soaked	24	1.10-1.30
	Dry	8	0.60-0.70
	Moist to wet	8	1.20-2.40
	Dry	24	0.20-0.40
	Moist	24	0.40-0.50
	Wet	24	0.60-0.90
	-	No soil cover	2-3
	Still	60 in. water plus	10
	River current	60 in. soil	2.0-2.5

After Karge (1945).

Table 5.2 Specific Heat Factor of 0.7 Specific Gravity Gas

Average Temperature in Flowline, °F	Flowline Pressure, psig										
	300	500	700	800	1000	1200	1500	1800	2100	2500	3000
120	29.1	30.3	31.0	31.6	32.5	33.3	34.8	36.2	37.2	28.8	40.6
100	28.7	29.9	30.8	31.4	32.4	33.4	35.1	36.7	38.0	39.7	41.6
80	28.2	29.5	30.5	31.3	32.4	33.5	35.4	37.2	38.7	40.5	42.5
60	27.5	29.2	30.3	31.1	-	-	-	-	-	-	-

After National Tank Company (1958).

Table 5.3 Average Ground Temperatures (T_g), °F

Cover, in.	T_g , °F
36	53 to 58
18	25 to 45 (Northern Europe, Canada, Alaska)
	45 to 48 (Northern United States, China, Russia)
	48 to 53 (Southern United States, Southeast Asia, West Africa, South America)

➤

5.6 TEMPERATURE DROP DETERMINATION

5.6.1 Overview

Choking (expansion of gas from high pressure to low pressure) is often required to control gas flow rates. Chokes and control valves are commonly used. Pressure drop across the restriction causes a decrease in gas temperature. If the gas is saturated with water and the final temperature of the gas is below the hydrate formation temperature, then hydrates will form. Pressure drop across a choke is a constant enthalpy process. For a multicomponent stream, one must perform flash calculations that balance enthalpy before and after the choke. This is best suited for a computer.

5.6.2 Temperature Drop Correlation (Figure 5.7)

The temperature drop correlation shown in Figure 5.7 can be used when the gas composition is unknown and for “quick look” or “first approximations.” It yields reliable results but is affected by liquids and thus adjustments must be made to correct for hydrocarbon liquids. The accuracy is $\pm 5\%$.

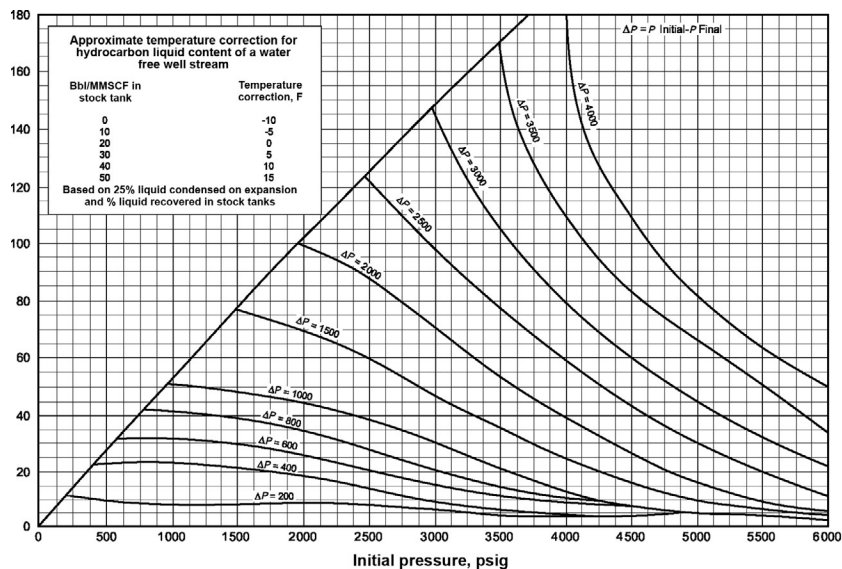


Figure 5.7 Temperature drop accompanying a given pressure drop for a natural gas stream.

Example 5.2: Determine the Temperature Drop Across a Choke

Given

A well with a flowing tubing pressure of 4000 psi and 20 bbl of hydrocarbon condensate/MMscfd with a downstream back pressure of 1000 psi.

Determine

Temperature drop across the choke.

Solution

Initial pressure = 4000 psi

Final pressure = 1000 psi

$\Delta P = 3000 \text{ psi}$

From [Figure 5.7](#) correlation; intersect initial pressure of 4000 psi and ΔP -3000 and read $\Delta T = 80^\circ \text{F}$. Note: If the amount of liquid hydrocarbons was more or < 20 bbl, a correction would need to be made that would affect the ΔT .



5.7 HYDRATE PREDICTION CORRELATIONS

5.7.1 Overview

All correlations are based on a system that contains only gas and water in a static test cell that was rocked only to provide good equilibrium. Data shown are the hydrate melting conditions, not the formation point. The correlations yield acceptable results. These correlations can be used to predict hydrate formation temperature.

5.7.2 Pressure-Temperature Curves

The results of using the pressure-temperature curves are not as accurate as vapor-solid equilibrium constants. They are used when the composition of the flow stream is not known.

They are typically used for “first approximations” or “quick look.”

5.7.3 Equations of State Calculations

Computer solutions have been developed to predict hydrate formation conditions.

5.7.4 Vapor-Solid Equilibrium Constants

Vapor-solid equilibrium constants yield reliable results up to 1000 psia. They can be used when the composition of the flow stream is known. The procedure to determine the hydrate formation temperature is as follows:

1. Assume hydrate formation temperature
2. Determine the equilibrium constant, K , for each component where

$$K_i = \frac{Y_i}{X_i} \quad (5.4)$$

where Y_i = mol fraction of each component in the gas on a water-free basis;
 X_i = mol fraction of each component in the solid on a water-free basis.

3. Calculate the ratio, Y_i/K_i , for each component
4. Sum the values of Y_i/K_i
5. Repeat steps 1–4 for additional temperatures until $\sum Y_i/K_i = 1$

Figures 5.8–5.12 are graphs giving vapor-solid equilibrium constants, K , at various pressure and temperatures.

Gas streams containing more than 30% H_2S behave as if they contain pure H_2S . Components heavier than butane have $K_i = \text{infinity}$, since their molecules are too large to fit into the cavities of the lattice structure.

Example 5.3: Determination of Hydrate Formation Temperature Using Vapor-Solid Constants

Given

A flow stream with a flowing pressure of 400 psia and the following composition.

Determine

The hydrate formation temperature.

Component	Mole Fraction of Gas
Nitrogen	0.0144
Carbon dioxide	0.0403
Hydrogen sulfide	0.000019
Methane	0.8555
Ethane	0.0574
Propane	0.0179
Isobutane	0.0041
<i>n</i> -Butane	0.0041
Pentane +	0.0063
	1.00000

Solution

Calculation of Temperature for Hydrate Formation at 400 psia

Component	Mole Fraction in Gas	At 70 °F		At 80 °F	
		K_i	Y_i/K_i	K_i	Y_i/K_i
Nitrogen	0.0144	Infinity	0.00	Infinity	0.00
Carbon dioxide	0.0403	Infinity	0.00	Infinity	0.00
Hydrogen sulfide	0.000019	0.3			
Isobutane	0.0041	0.15	0.03	0.06	0.01
<i>n</i> -Butane	0.0041	0.72	0.00	1.22	0.00
Pentane +	0.0063	Infinity	0.00	Infinity	0.00
Total	1.0000		1.08		0.87

Note:

- (1) Interpolating linearly, V/K : 1.0 at 74 °F. Therefore, hydrates will form at 75 °F.

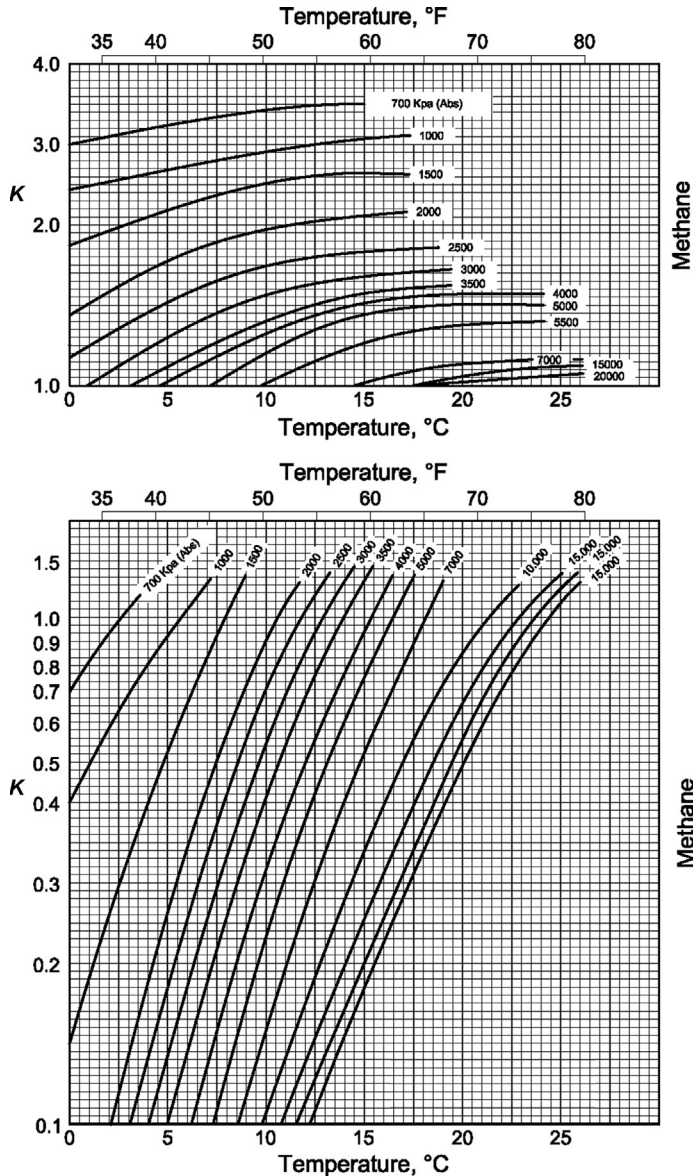


Figure 5.8 Vapor-solid " K " values for methane and ethane. (Adapted to SI by GPSA. Original courtesy of AIIME).

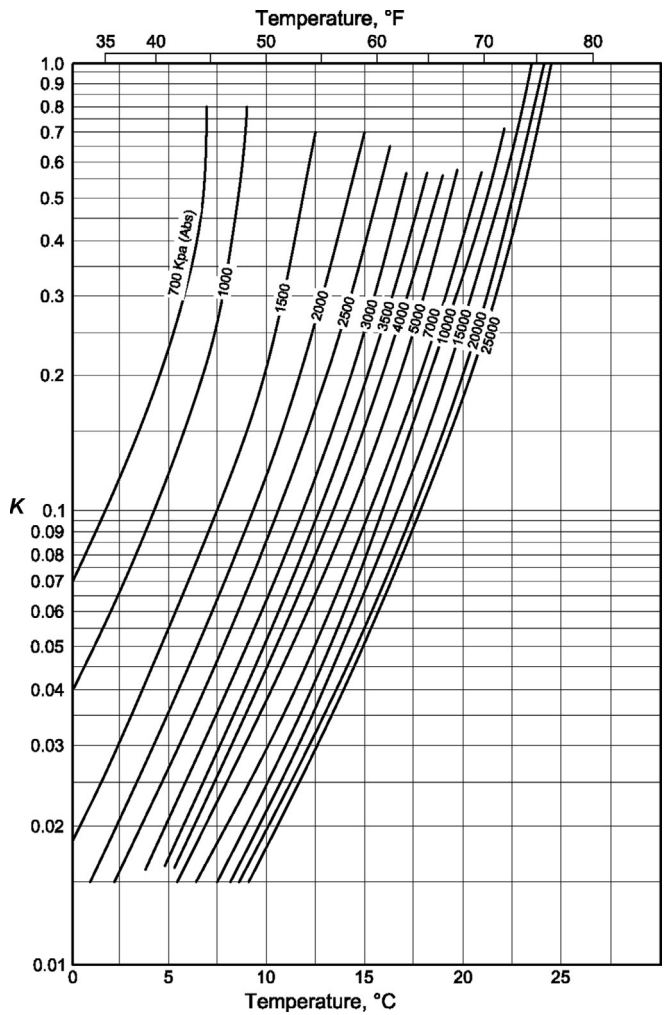


Figure 5.9 Vapor-solid “K” values for propane. (Adapted to SI by GPSA. Original courtesy of AIME).

5.7.5 Pressure-Temperature Curves (Figure 5.13)

Pressure-temperature curves are used when gas composition is not known or for “first approximation.” Graphs have been developed to approximate hydrate formation temperature as a function of gas gravity and pressure.

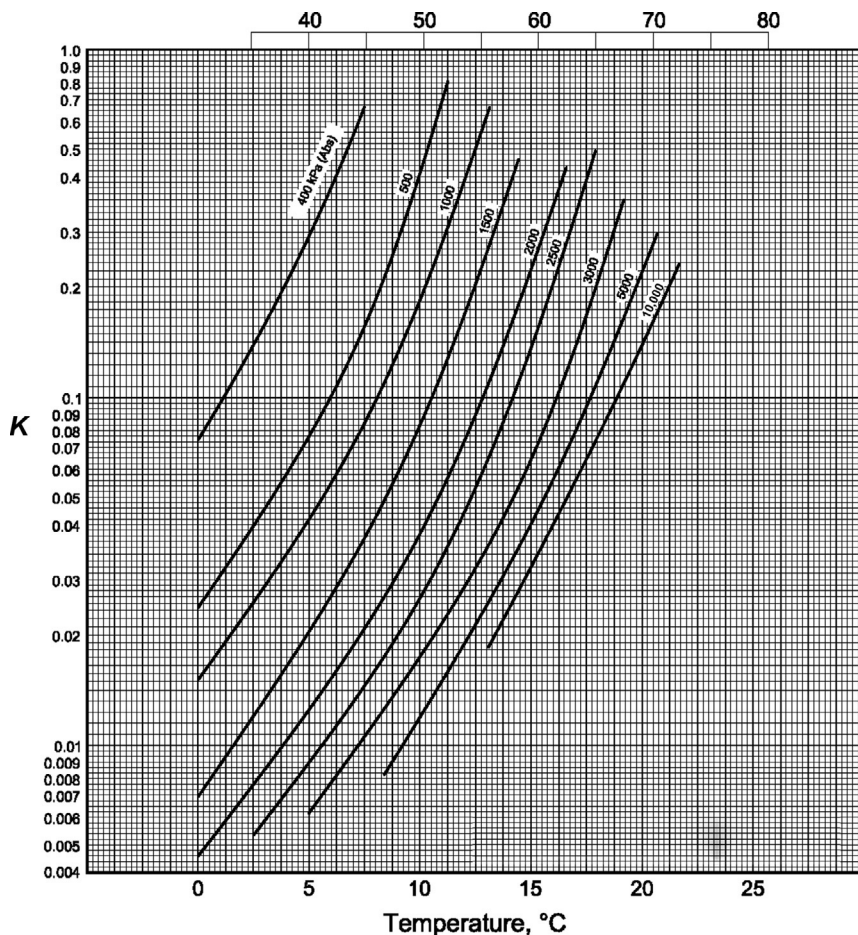


Figure 5.10 Vapor-solid “K” values for isobutane. (Adapted to SI by GPSA. Original courtesy of AIIME).

Example 5.4: Determine the Hydrate Formation Temperature Using Pressure-Temperature Correlations

Given

A 0.6 specific gravity gas operating at 2000 psia.

Determine

Calculate the Hydrate Formation Temperature using the Pressure-Temperature Correlations.

Solution

From pressure-temperature curve intersect 2000 psia and 0.6 specific gravity curve and read 68 °F.

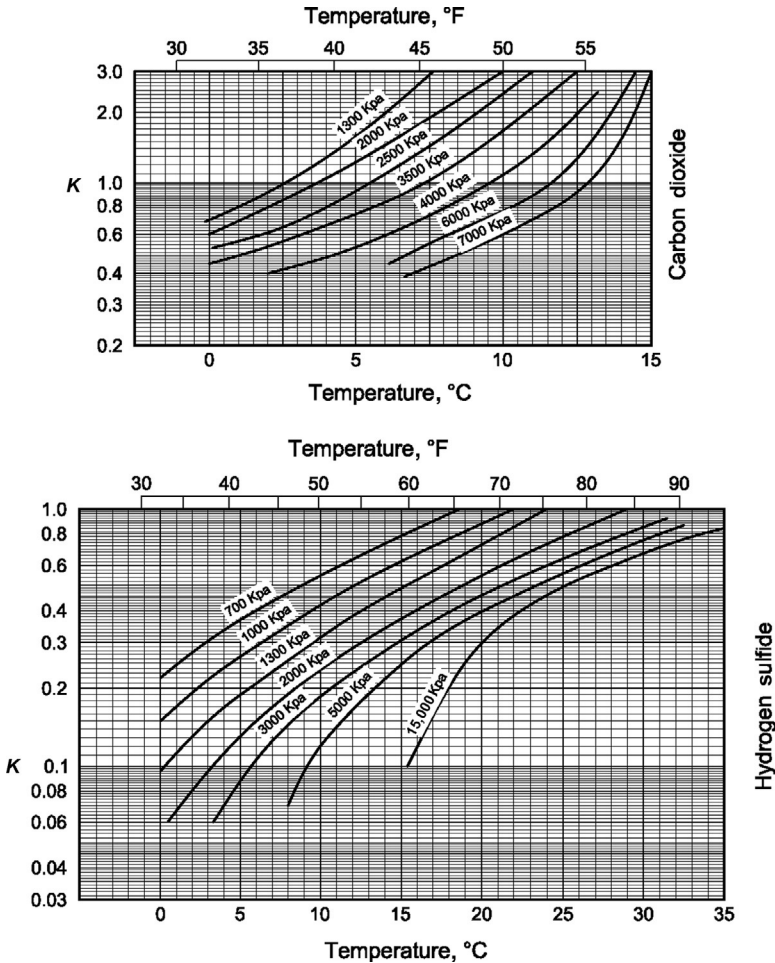


Figure 5.11 Vapor-solid “ K ” values for CO₂ and H₂S. (Adapted to SI by GPSA. Original courtesy of AIIME).



5.8 HYDRATE PREVENTION

5.8.1 Overview

Hydrate prevention is used to prevent hydrates from forming. Operating conditions must remain out of the hydrate-formation zone. Hydrate point must be maintained below the operating conditions of the system. Two common methods of hydrate-formation prevention are:

- Temperature control
- Chemical injection

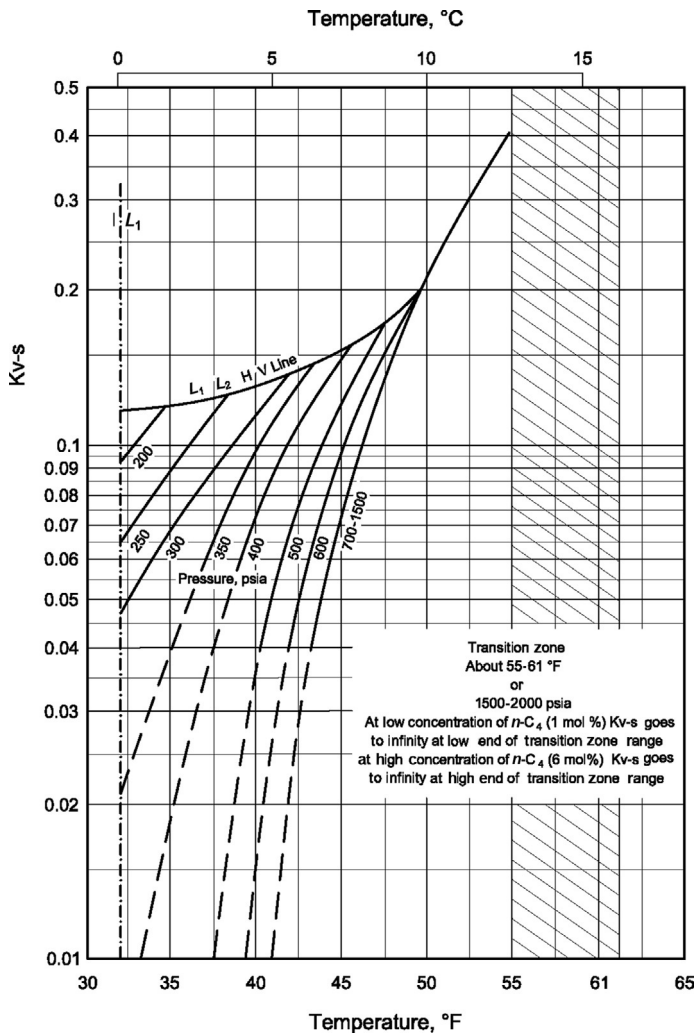


Figure 5.12 Vapor-solid values for normal-butane. (Adapted to SI by GPSA. Original courtesy of AIME).

5.8.2 Adding Heat

Adding heat is effective because hydrates normally do not occur above 70 $^{\circ}\text{F}$ (21 $^{\circ}\text{C}$). It offers a simple and economical solution for land and offshore facilities (if waste heat is available). Flow stream is preheated, either through an indirect line heater or heat exchanger, before passing through a choke. Flow stream is then reheated to maintain the temperature above the hydrate formation temperature.

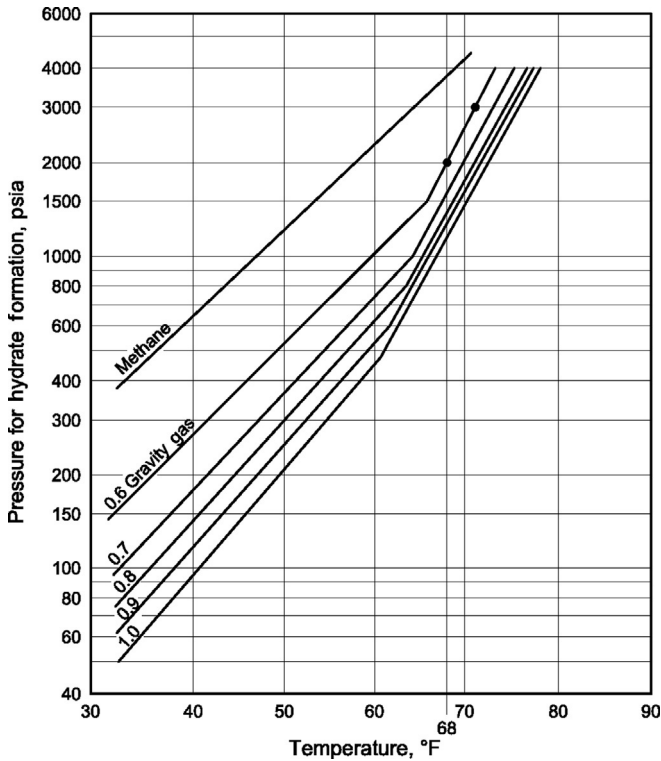


Figure 5.13 Pressure-temperature curves for predicting hydrate formation temperature.

A major drawback in offshore installations is that it is almost impossible to maintain flowline temperatures significantly above the water temperature if the flowlines extend more than a few hundred feet under water. Thus, either the “free water” must be separated while still at temperature, or an alternate method must be selected.

5.8.3 Temperature Control

5.8.3.1 Indirect Heaters

5.8.3.1.1 Overview

An indirect heater is used to heat gas to maintain temperatures above that of hydrate formation. It consists of an atmospheric vessel containing a fire tube (usually fired by gas, steam, or heating oil) and a coil (designed to withstand shut-in tubing pressure (SITP)), that is heated by the intermediate fluid (usually water) and the fluid is heated. The fire tube and coil are immersed in a heat transfer fluid (normally water), and heat is transferred to the fluid in the coil.

5.8.3.1.2 Wellhead Heater Description (Figures 5.14 and 5.15)

Figure 5.14 shows a typical heater installation at the wellhead.

Beginning at the wellhead, the following items are normally included:

5.8.3.1.2.1 Safety Shut-Down “Wing” Valve A pneumatically actuated valve that is connected directly to the Christmas tree PSL pilot, will shut-in the well whenever the flowline pressure upstream of the heater falls below a certain set pressure, indicative of a flowline rupture.

PSHL pilot, senses flowline pressure downstream of the heater choke and will shut-in the well on either abnormally high or low pressure.

5.8.3.1.2.2 High-Pressure Flowline A high-pressure flowline is line, normally at least 150 ft. long, designed to withstand full wellhead SITP.

5.8.3.1.2.3 Expansion Loop An expansion loop is designed to absorb flowline-length changes caused by changes in temperature between flowing and shut-in conditions.

5.8.3.1.2.4 Long-Nose Heater Choke (Figure 5.15) A long-body choke, also called a long-nose heater choke, is installed in the indirect heater to position the choke orifice within the indirect heater bath. Since the walls of the choke orifice are heated by the water bath, hydrates will not form in the orifice and cause plugging.

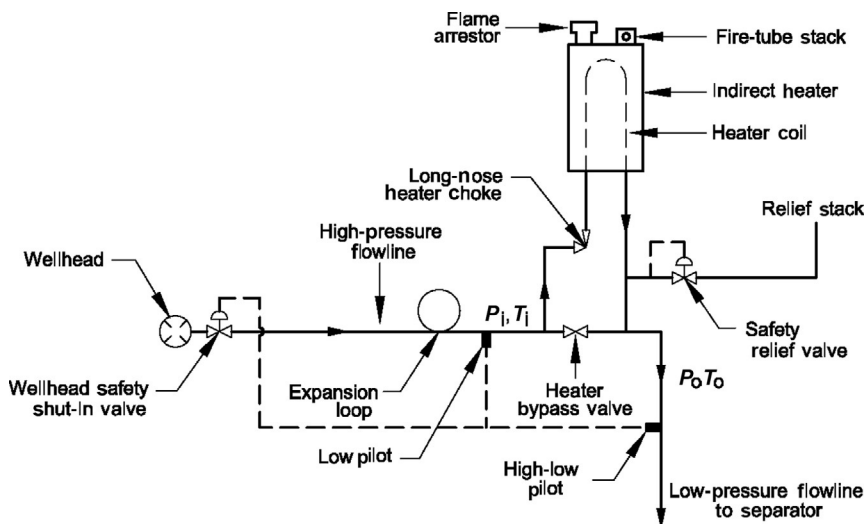


Figure 5.14 Typical wellhead indirect heater schematic.

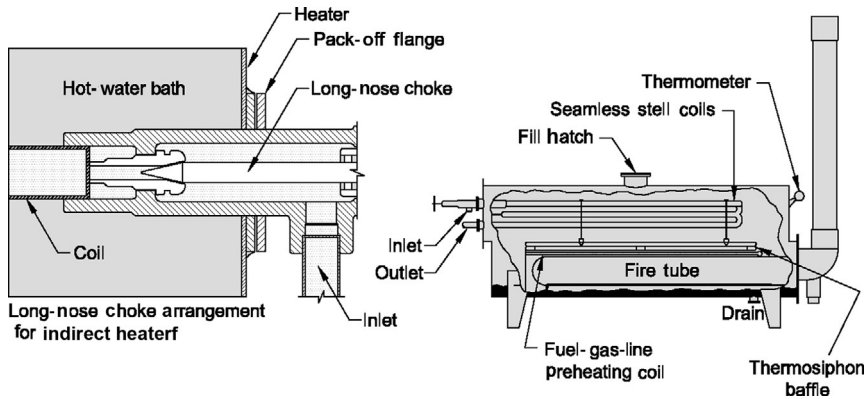


Figure 5.15 Indirect heater details.

5.8.3.1.2.5 Heater Bypass Valve The heater bypass valve is designed to withstand full wellhead shut-in pressure and bypass gas around the heater after the wellhead pressure has been drawn down to near-sales line pressure. Use of this valve prevents needless wear and erosion on the heater coils and allows the pressure drop from wellhead to sales line to be minimized.

5.8.3.1.2.6 Heater Coil A multiple-pass steel coil, or heater coil, is designed to withstand full wellhead SITP.

Because of high corrosion and erosion rates, the return bends are “safety-drilled.”

Holes will begin to leak when corrosion and erosion have reduced the wall thickness by half, warning that the bends should be replaced.

5.8.3.1.2.7 Pressure-Relief Valve A pressure-relief valve provides over-pressure protection for the low-pressure flowline.

5.8.3.1.2.8 Heater Flame Arrestor The heater flame arrestor is a device that provides fire protection by preventing the heater fire from flashing back through the air intake and igniting surrounding material.

5.8.3.2 Flowline Heaters

Flowline heaters differ from wellhead heaters in purpose only. The purpose of a wellhead heater is to heat the flow stream at or near the wellhead where choking or pressure reduction occurs.

The purpose of a flowline heater is to provide additional heat if required. The design is the same as an indirect heater, except the choke, shut-in, and relief equipment are seldom used. A bypass should be installed in either case so that the heater can be taken out of service.

5.8.3.3 System Optimization

System operation has to be optimized before heaters can be effectively designed and located.

Heat requirements that appear to be large often can be reduced to minimal values, or even eliminated, by revising the mode of operation. For example, fields having multiple producing wells can be combined to use higher flowing temperatures thus minimizing heater requirements.

If reducing the gas stream pressure is necessary, it is generally more efficient to do so at a central point where the necessary heater fuel gas can be obtained from separators or scrubbers. This requires flowline wall thickness to be increased to withstand wellhead SITP. An alternative is to install wellhead shut-down valves and flowline PSHs.

5.8.3.4 Heater Sizing

To adequately describe the size of a heater, heat-transfer duty and coil sizes must be specified.

To determine the heat duty required, one must know:

- Amount of gas, water, and oil or condensate expected
- Pressures and temperatures of the heater inlet and outlet

The heater outlet temperature depends on the temperature at which hydrates form.

The coil size depends on the volume of fluid flowing through the coil and the required heat-transfer duty. Special operating conditions such as start-up of a shut-in well should be considered.

5.8.3.5 Downhole Regulators

Downhole regulators are feasible for high-capacity gas wells at locations where certain risks to other downhole equipment are acceptable. The theory behind the use of a downhole regulator is that the pressure drop from flowing pressure to near-sales line pressure is taken downhole where the formation temperature is sufficiently high to prevent hydrate formation. The tubing string above the regulator then acts as a subsurface heater. Calculation of heat duty and sizing procedures are discussed in detail in [Chapters 3 and 4](#).

Calculations involved in downhole regulator design are rather complicated. They depend on characteristics such as: wellbore configuration, flowing downhole pressures and temperature, and well depth. Although shortcut procedures are available to estimate the feasibility of downhole regulators, tool company representatives can provide detailed design information.

5.8.4 Chemical Injection

5.8.4.1 Overview

Hydrate inhibitors are used to lower the hydrate formation temperature of the gas. Methanol and ethylene glycol are the most commonly used inhibitors. Recovery and regeneration steps are used in all continuous glycol injection projects and in several large-capacity methanol injection units. Injection of hydrate inhibitors should be considered for the following applications:

- Pipeline systems in which hydrate trouble is of short duration
- Gas pipelines that operate at a few degrees below the hydrate formation temperature
- Gas-gathering systems in pressure-declining fields
- Gas lines in which hydrates form as localized points

Methanol and the lower molecular weight glycols have the most desirable characteristics for use as hydrate inhibitors.

Table 5.4 lists some physical properties of methanol and the lower molecular weight glycols.

When hydrate inhibitors are injected in gas flowlines or gathering systems, installation of a free-water knockout at the wellhead proves to be economical in nearly every case. Removing the free water from the gas stream reduces the amount of inhibitor required.

5.8.4.2 Methanol Injection Considerations

Methanol is well-suited for use as a hydrate inhibitor because it is:

- Noncorrosive
- Nonreactive chemically with any constituent of the gas
- Soluble in all proportions in water
- Volatile under pipeline conditions
- Reasonable in cost
- Has a vapor pressure greater than that of water

5.8.4.3 Methanol Injection System Description (Figure 5.16)

Methanol is injected by means of a gas-driven pump, (3) in Figure 5.16, into the flowline upstream of the choke or pressure control valve (2). A temperature controller (5) measures the temperature of the gas in the low-pressure flowline (7) and adjusts the methanol rate accordingly.

The methanol injection rate is controlled by the amount of power gas allowed to flow through the power gas control valve (4) to drive the pump.

Table 5.4 Physical Properties of Chemical Inhibitors

Property	Methanol	Ethylene Glycol	Diethylene Glycol	Triethylene Glycol	Tetraethylene Glycol
Molecular weight	32.04	62.10	106.10	150.20	194.23
Boiling point at 760 mmHg, °F	148.10	387.10	427.60	532.90	597.2
Vapor pressure at 77 °F, mmHg	94	0.12	<0.01	<0.01	<0.01
Specific gravity at 77 °F	0.7868	1.110	1.113	1.119	1.120
Specific gravity at 140 °F	-	1.085	1.088	1.092	1.092
Pounds per gallon at 77 °F	6.55	9.26	9.29	9.34	9.34
Freezing point, °F	-144	8	17	19	22
Pour point, °F	-	<-75	-65	-73	-42
Absolute viscosity in centipoises at 77 °F	0.55	16.5	28.2	37.3	39.9
Absolute viscosity in centipoises at 140 °F	0.36	5.1	7.6	9.6	10.2
Surface tension at 77 °F, dynes/cm	22	47	44	45	45
Specific heat at 7 °F, Btu/lb/°F	0.27	0.58	0.55	0.53	0.52
Flash point, °F	0	240	280	320	365
Fire point, °F	0	245	290	330	375
Decomposition temperature, °F	0	329	328	404	460
Heat of vaporization at 14.65 psi, Btu/lb	473	364	232	179	-

5.8.4.4 Glycol Injection Considerations

Glycol has a relatively low vapor pressure and thus does not evaporate into the vapor phase as readily as methanol. The solubility of glycol in liquid hydrocarbons is relatively low.

For the above reasons, glycol can be more economically recovered, thus reducing the operating expenses below those of methanol systems.

5.8.4.5 Glycol Injection and Recovery System Description (*Figure 5.17*)

The injection part of the system (items 1 to 5 in *Figure 5.17*) is similar to the methanol injection system. Additional equipment in the glycol system is for recovering and reclaiming the glycol.

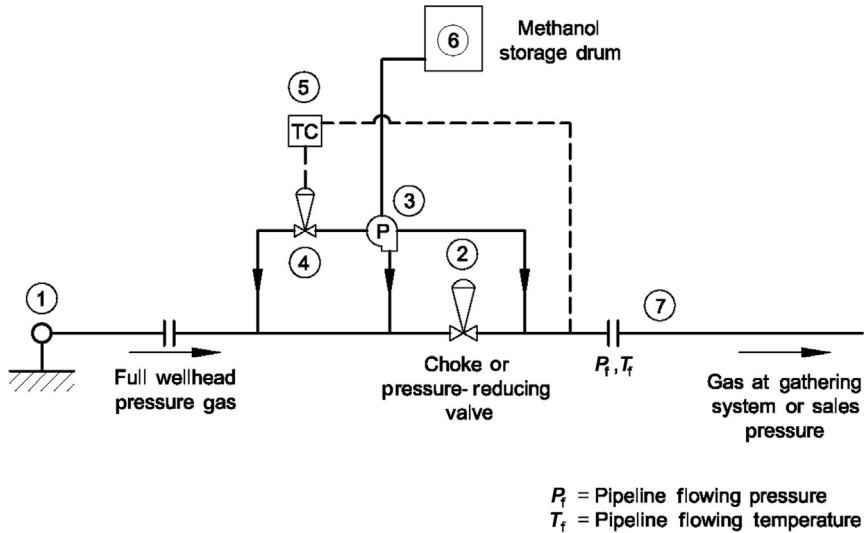


Figure 5.16 Typical methanol injection system.

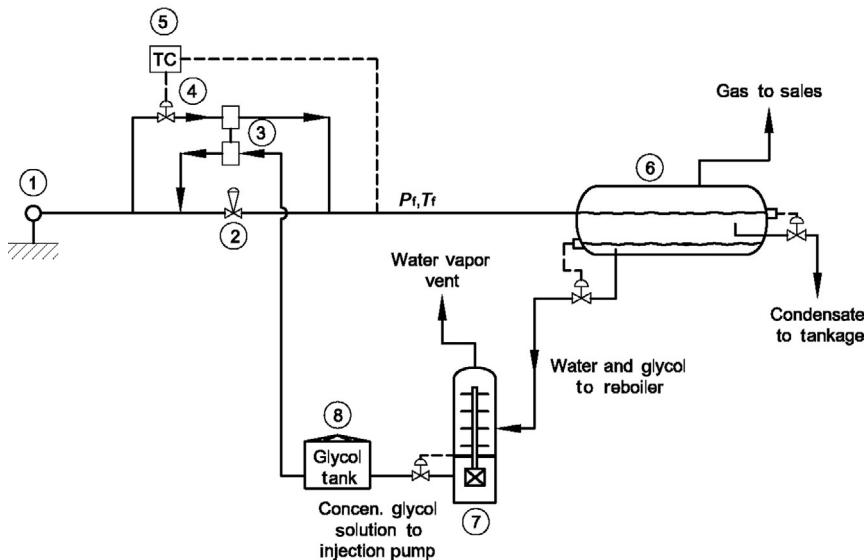


Figure 5.17 Typical glycol injection and recovery system.

A three-phase separator (6) separates the water and glycol from the hydrocarbon phases. The water-glycol solution in the separator is sent to the reboiler (7) while gas is delivered to the sales line and the hydrocarbon condensate is dumped to the condensate storage tanks. In the reboiler, excess water is boiled away from the glycol. The glycol reconcentrated in the

reboiler is then available again for injection into the gas stream. Separation of the glycol water phase from the hydrocarbon-liquid field requires a temperature above 70 °F (20 °C) and a residence time of 10–15 min.

5.8.4.6 Nozzle Design (Figure 5.18)

Due to the vapor pressure of glycol, a fine, well-distributed mist is required to obtain adequate mixing with the gas to ensure optimum results, thus spray nozzles are normally used (refer to Figure 5.18). Nozzle selection is a major consideration in the design of cold-separation facilities or plants using glycol injection. Glycol injection normally takes place just upstream of a heat exchanger or chiller where gas is being chilled. Proper nozzle selection will ensure that the glycol spray covers the tube sheet. To atomize the glycol, 100–150 psi differential pressure at the nozzle is sufficient. Process stream velocities should be at least 12 fps.

5.8.4.7 Glycol Selection

The three glycols typically used to prevent the formation of hydrates are:

- Ethylene glycol (EG)
- Diethylene glycol (DEG)
- Triethylene glycol (TEG)

Selection of a glycol depends on the composition of the hydrocarbon flow stream and the advice of the glycol supplier.

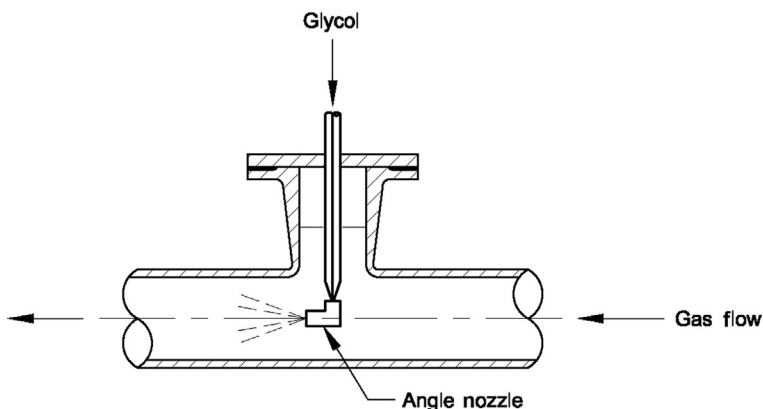


Figure 5.18 Schematic of a spray nozzle used in glycol injection.

5.8.4.8 Glycol Injection Guidelines

If glycol is to be injected into a natural gas transmission line where glycol recovery is of less importance than hydrate protection, ethylene glycol is the best choice because it produces the greatest hydrate depression and has the highest vapor pressure of any of the glycols. If glycol is to be injected into a unit where it will contact hydrocarbon liquids, ethylene glycol is preferred because it has the lowest solubility in high molecular weight hydrocarbons.

If vaporization losses are severe, either diethylene or triethylene glycol is a better choice because both have a lower vapor pressure. Sometimes, diethylene glycol is used if there is a combined loss of both gas vaporization and liquid solubility.

The freezing point of the glycol solution must be lower than the lowest temperature expected in the system. In inhibitor service, glycol concentrations are usually maintained at 70–75 wt% because freezing of the glycol is not a problem at this concentration.

Reboiler temperature is dependent on the type of glycol and its concentration. Temperature should be maintained at a level equal to the boiling point of the desired solution. Boiling points for the three glycol types are plotted in [Figures 5.19, 5.20, and 5.21](#). For example, from [Figure 5.19](#) the reboiler temperature should be set at 240 °F (116 °C) to produce a 70 wt% ethylene glycol solution at atmospheric pressure (760 mm). Thermal degradation can occur if the boiling point of the pure glycol is exceeded; it should therefore be avoided. Glycol losses for the two-phase gas-condensate systems are normally estimated at 1 to 2 gallons per 100 barrels of hydrocarbon liquid produced.

Vaporization into the gas stream and solution into the hydrocarbon liquid usually cause only a small portion of the total loss. The most significant causes of glycol losses are leakage and carryover with the hydrocarbon liquid. Losses also occur from vaporization and carryover in the reboiler.

5.8.4.9 Injection Requirement Categories

5.8.4.9.1 Low Pressure-High Volume

Pressures up to 2000 psi and volumes measured in hundreds to thousands of BPD.

5.8.4.9.2 High Pressure-Low Volume

Pressures up to 15,000 psi and volumes measured in quarts or a few gph.

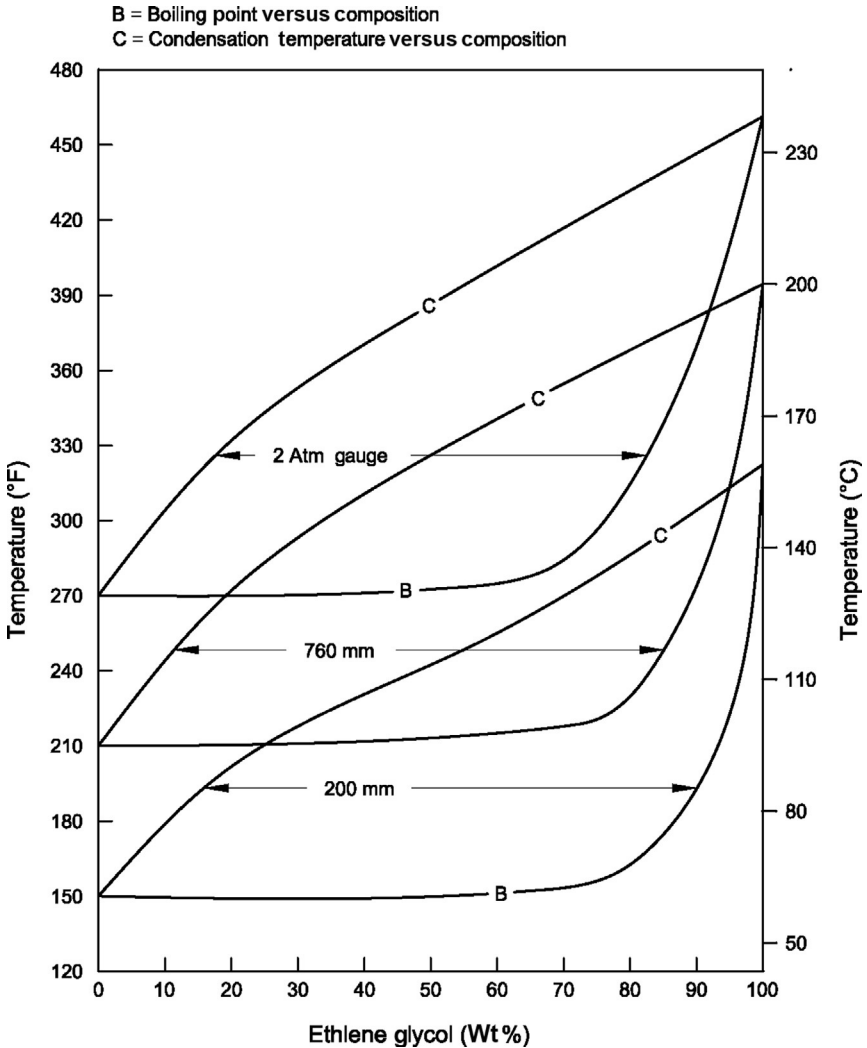


Figure 5.19 Boiling points and condensation temperatures of aqueous ethylene glycol solutions at various pressures.

5.8.4.9.3 High Pressure-High Volume

High pressure-high-volume cases are the most difficult to deal with, and describe pressures exceeding 5000 psi and volumes measured in several gpm or barrels per minute. Intermittent control when problems are encountered as “local” heat can be applied at the surface to remove hydrate plugs that may occur.

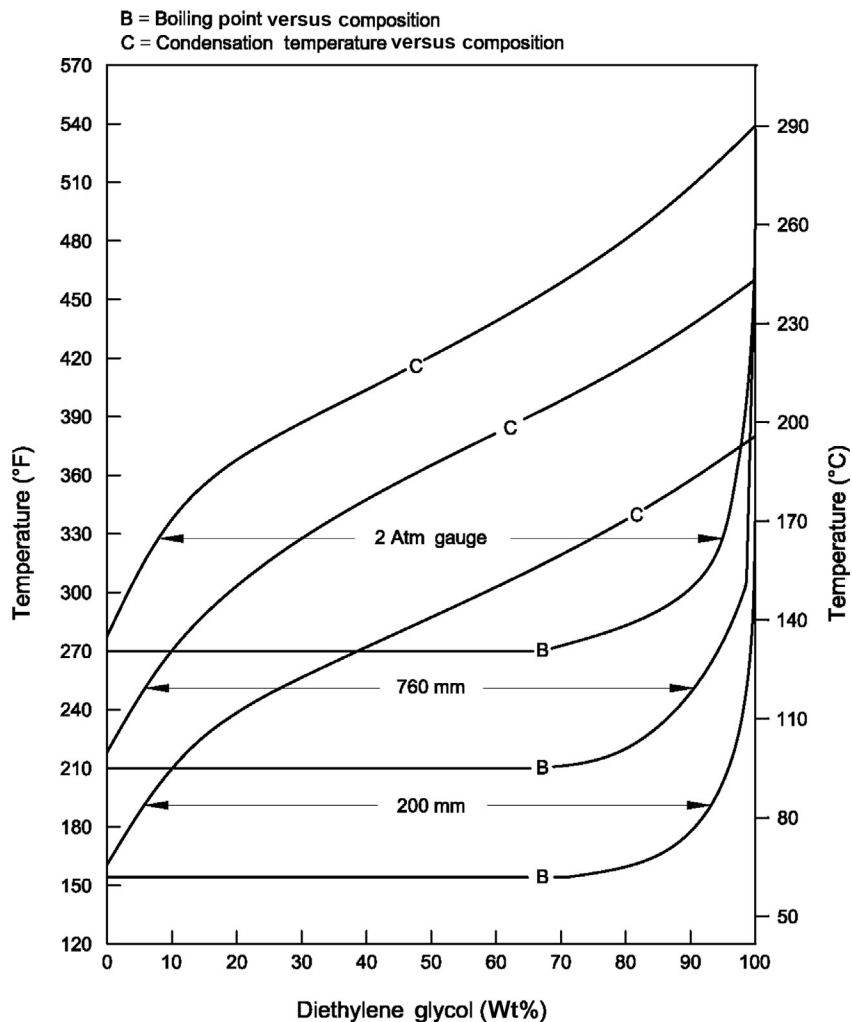


Figure 5.20 Boiling points and condensation temperatures of aqueous diethylene glycol solutions at various pressures.

5.8.4.10 Single versus Two-Step Injection Considerations

5.8.4.10.1 Single-Step Injection

All chemicals are injected through a downhole tubing mandrel that handles both wellhead and flowline conditions.

5.8.4.10.2 Two-Step Injection

Two-step injection utilizes a second injection point just downstream of the wellhead to handle additional water that condenses from the gas phase as the flow stream cools to the surrounding temperature.

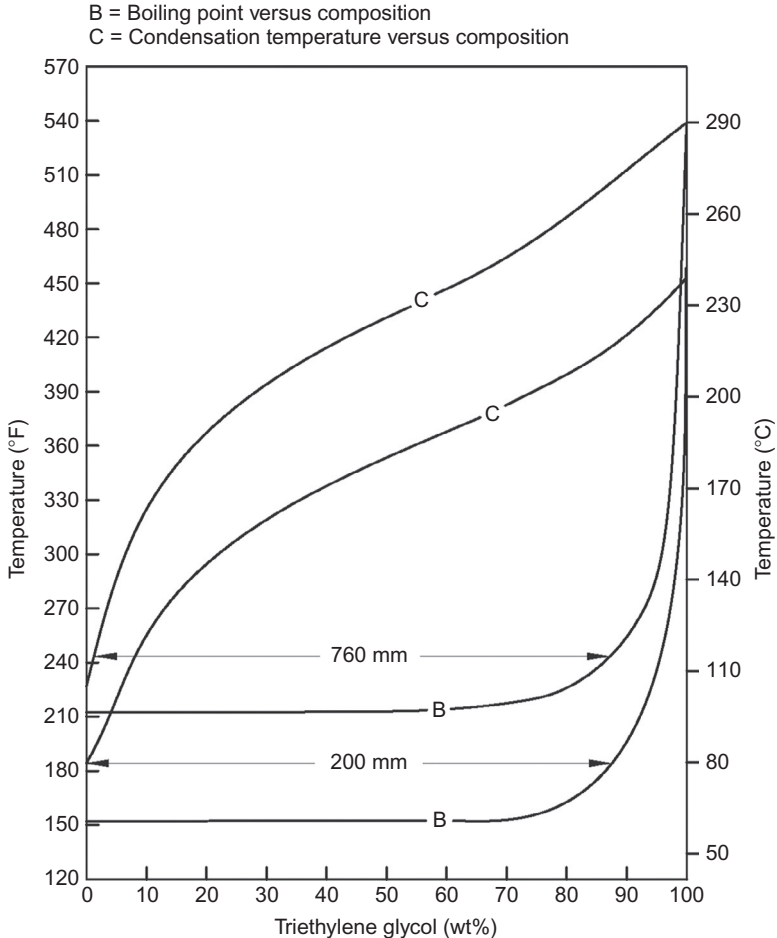


Figure 5.21 Boiling points and condensation temperatures of aqueous triethylene glycol solutions at various pressures.

5.8.5 Chemical Injection System

5.8.5.1 Overview

The three parts of an injection system are the pump, meter, and control system.

5.8.5.2 Single-Point Chemical Injection

A single-point chemical injection system has pump, meter, and control system service at one injection point.

Disadvantages include:

- Limited turndown capability and increased life-cycle cost
- Weight and space increase as injection points increase

5.8.5.3 Multi-Point Chemical Injection

A multi-point chemical injection system consists of a shared pump and multiple meter and control devices servicing multiple injection points.

Advantages

- Increased turndown capacity and inferred flow monitoring
- Self-compensated with closed loop control
- Per well capital investment decreases as the number of wells increases
- Injection points are easily added
- Lower weight and space requirements for higher quantity well applications

Disadvantages

- Instrumentation intensive
- Requires multiple control loops
- Requires variable speed for fixed crank pumps
- Experiences high-pressure drops from header to recycle line

5.8.5.4 Metering Pump Considerations

When selecting a metering pump, the following should be considered (refer to API 675):

- Features of pump-meter-control functions
- Vertical or horizontal
- Variable crank
- Robust design
- Modular construction
- Pumphead interchangeability
- Highly precise and repeatability

5.8.5.5 Diaphragm Pumps

Advantages

- Hermetically sealed, no contamination to atmosphere
- Long-life diaphragms typically greater than 2 years continuous duty (20,000 hours).
- Long-life of hydraulic plunger seals, typically >2 years continuous duty (20,000 h)
- Internal hydraulic relief
- Maximum safeguard to environment and personnel safety automated diaphragm failure mechanisms

Disadvantages

- Higher purchase price (pay back in less downtime)
- More complex maintenance required

5.8.5.6 *Plunger Pumps*

Advantages

- Lower purchase price
- Less complicated maintenance (easier to understand)

Disadvantages

- Plunger packing service life is typically, <2000 h
- Friction between plunger and packing

5.8.6 Comparison of Hydrate Prevention Methods

5.8.6.1 *Overview*

The four methods (indirect heaters, methanol injection, glycol injection, and downhole regulators) discussed above are proven safe and reliable. Evaluation should consider:

- Development of CAPEX and OPEX (including chemicals and fuel)
- Space needs (especially in offshore operations) and operating hazards

5.8.6.2 *Heaters*

Capital costs and the fuel expense of heaters are relatively large, and it is difficult to maintain a clean, reliable fuel supply to remote heater locations. Indirect heaters require a large amount of space. Fire boxes with proper flame arrestors have minimized the hazards from fired equipment, but they should be bought with strict attention paid to detailed design.

5.8.6.3 *Chemical Injection*

Advantages and disadvantages of methanol injection and glycol injection are listed in [Table 5.5](#).

Table 5.5 Methanol and Glycol Injection Comparisons

Inhibitor	Advantages	Disadvantages
Methanol	Relatively low initial cost	High operating cost
	Minimal equipment	Hauling to site necessary
	Simple system with little gas consumption	
Glycol	Usually lower operating cost than methanol when both systems recover chemical	High initial cost
		Hauling to site necessary
	Simple system with little gas consumption	Large loss if line breaks
		Possibility of glycol concentration

The use of methanol requires only a free-water separator and a suitable means for injection and atomizer, whereas the use of glycol requires a free-water separator plus a gas-liquid separator and a glycol re-concentration unit at the point of recovery downstream.

5.8.6.4 Downhole Regulators

Downhole regulators do not require routine service, but a wireline service company must be used each time the pressure drop has to be changed and when the regulator is removed. A well with a downhole regulator may require injection of methanol or glycol when it is brought back online after a shut-in until the flow and temperature stabilize. After a well declines to less than allowable production, the downhole regulator will have to be removed and another form of hydrate prevention may prove necessary. Downhole regulators do not present special safety hazards, but because work with regulators involves working in the well, losing the well is always a danger.

5.8.7 Summary of Hydrate Prevention Methods

The methanol injection system is often used for temporary hydrate prevention service in small installations. Larger installations are favored for indirect heaters or glycol injection systems.

Downhole regulators are most useful in large high-pressure reservoirs in which excess pressure is available and the reservoir pressure is not expected to decline rapidly. Table 5.6 contains a summary comparison of the above methods.



5.9 HYDRATE INHIBITION

5.9.1 Hammerschmidt Equation

The Hammerschmidt equation is used to determine the amount of inhibitor required in the water phase to lower the hydrate temperature. It is expressed as:

$$\Delta T = \frac{KW}{100(MW) - (MW)(W)} \quad (5.5)$$

where ΔT = depression of hydrate formation temperature, °F;
 MW = molecular weight of inhibitor; K = constant, from table below;
 W = weight percent of inhibitor in final water.

Table 5.6 Comparison of Hydrate Prevention Methods

Technique	Investment	Fuel	Operating Maintenance	Chemicals	Plot Area	Hazards	Downtime
Down hole regulators	Very low	None	None	None	None	High	Low
Wellhead heaters	Very high	Very high	Low	Very low	Very high	High	Low
Methanol injection	Very low	None	Low	Very high	Very low	Medium	Low
Glycol injection	High	Medium	Low	High	Very high	High	Low

Inhibitor	Constants	
	<i>MW</i>	<i>K</i>
Methanol	32.04	2335
Ethanol	46.07	2335
Isopropanol	60.10	2335
Ethylene glycol	62.07	2200
Propylene glycol	76.10	3540
Diethylene glycol	106.10	4370

5.9.2 Determination of Total Inhibitor Required

$$\left(\begin{array}{c} \text{Total} \\ \text{Inhibitor} \\ \text{Required} \end{array} \right) = \left(\begin{array}{c} \text{Inhibitor} \\ \text{required} \\ \text{freewater} \end{array} \right) + \left(\begin{array}{c} \text{Inhibitor} \\ \text{lost to} \\ \text{vapor phase} \end{array} \right) + \left(\begin{array}{c} \text{Inhibitor} \\ \text{soluble} \\ \text{condensate} \end{array} \right) \quad (5.6)$$

where inhibitor lost to vapor phase is determined from [Figure 5.24](#). Methanol lost to the vapor phase. The inhibitor soluble in the condensate is approximately 0.5%.

5.9.3 Procedure for Determining Inhibitor Requirements

The procedure is best illustrated by an example; refer to [Figure 5.22](#).

Example 5.5: Determining the Amount of Methanol Required in a Wet Gas Stream

Given

FWHT = 100 °F for the subsea well

Determine

Calculate the total methanol required to prevent hydrates from forming. A conservative approach is to assume that the gas is saturated at wellhead conditions.

Solution

1. The amount of water that will be condensed is determined from McKetta-Wehe ([Figure 5.23](#)), assuming the gas is saturated at reservoir and wellhead conditions.

Water Content = 32.0 lb/MMSCF at (3000 psia and 100 °F) at wellhead

Water Content = −11.5 lb/MMSCF at (2000 psia and 60 °F) at platform

Subsea well tie-back to an existing shallow water production platform

Flowrate : (Multi-phase)

Gas = 20 MMSCFD (SG=0.600)

Condensate = 800 bcpd (60°API/SG=0.739)=bbl/MMSCF

Prod. water = 60 bwpd (SG=1.03) = 3 bbl/MMSCF

SIWHP = 3800 psia

FWHP = 3000 (P_1)

FWHT = 100 °F (T_1)

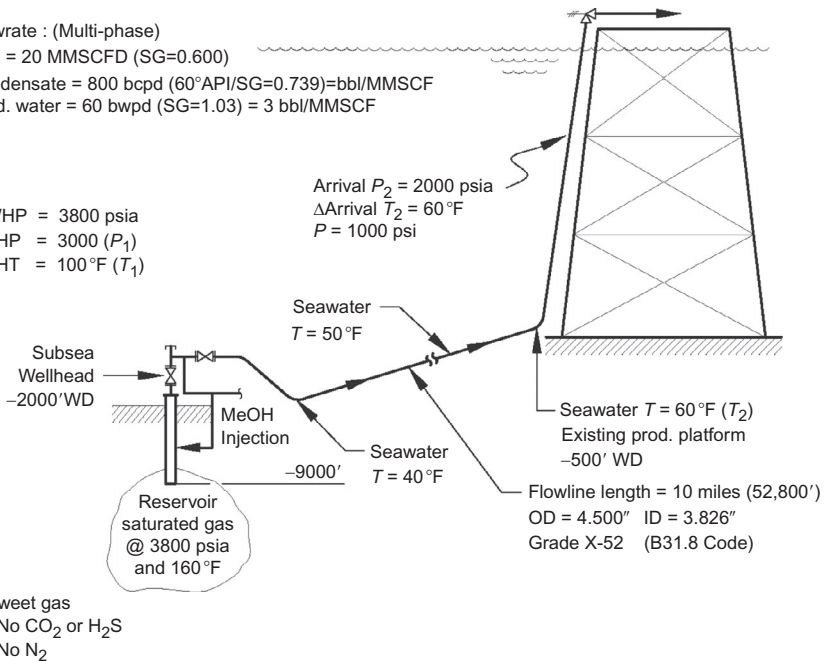


Figure 5.22 Subsea methanol injection example.

Water Condensed = 20.5 lb/MMSCF

Produced Water = 1083 lb/MMSCF

Total = 1103.6 lb/MMSCF

- From pressure-temperature curve, the hydrate formation temperature is 68 F (refer to Figure 5.13).

The required dew-point depression then is $68 - 60$ °F = 8 °F.

- The concentration of methanol required in the liquid water phase from Equation (5.5) is

$$8^\circ\text{F} = \frac{2335W}{100(32.042) - (32.042)W}$$

Rearranging and solving for $W = 9.892\% = 0.09892$.

- Therefore, the estimated methanol required in the liquid water phase is:

$$= \frac{0.09892}{1 - 0.09892} \left(1103.65 \frac{\text{lb}}{1 \text{ MMSCF}} \right)$$

$$= 121.15 \text{ lb/MMSCF}$$

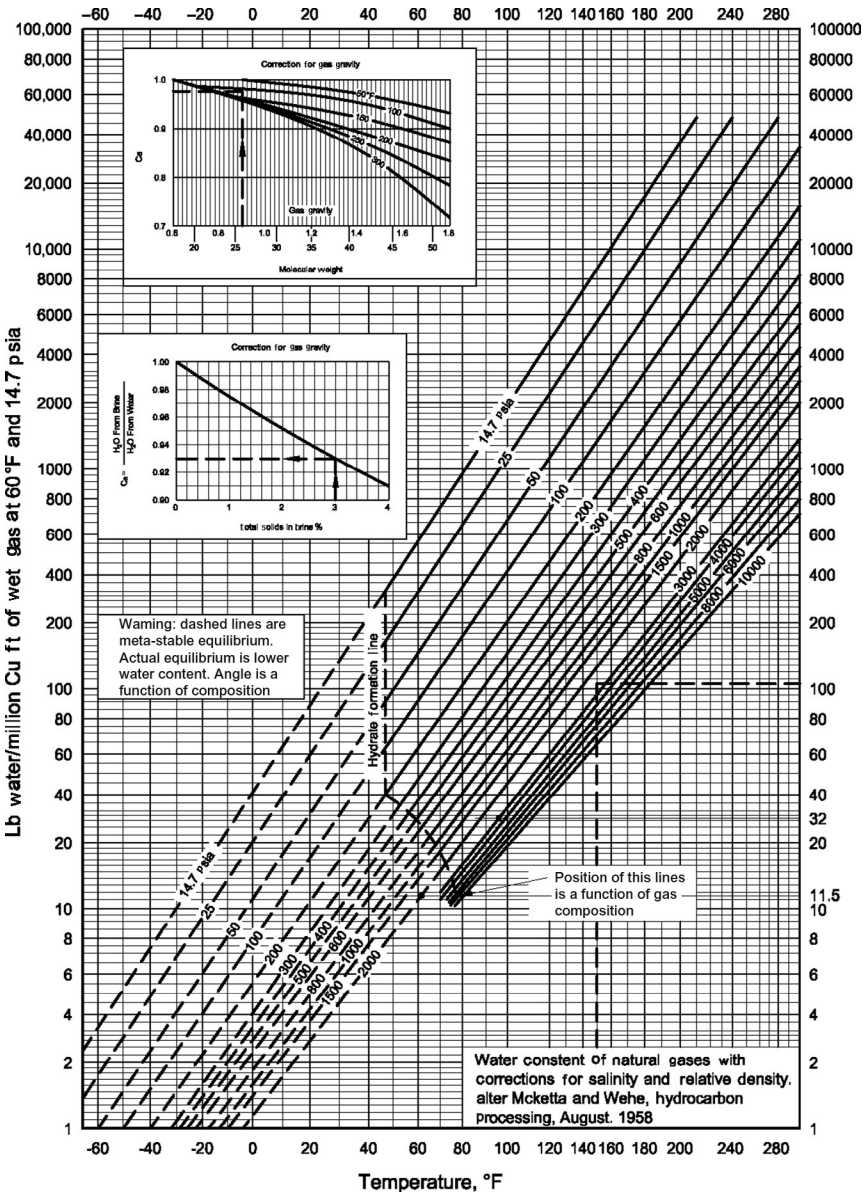


Figure 5.23 Water content of sweet, lean natural gas—McKetta-Wehe example.

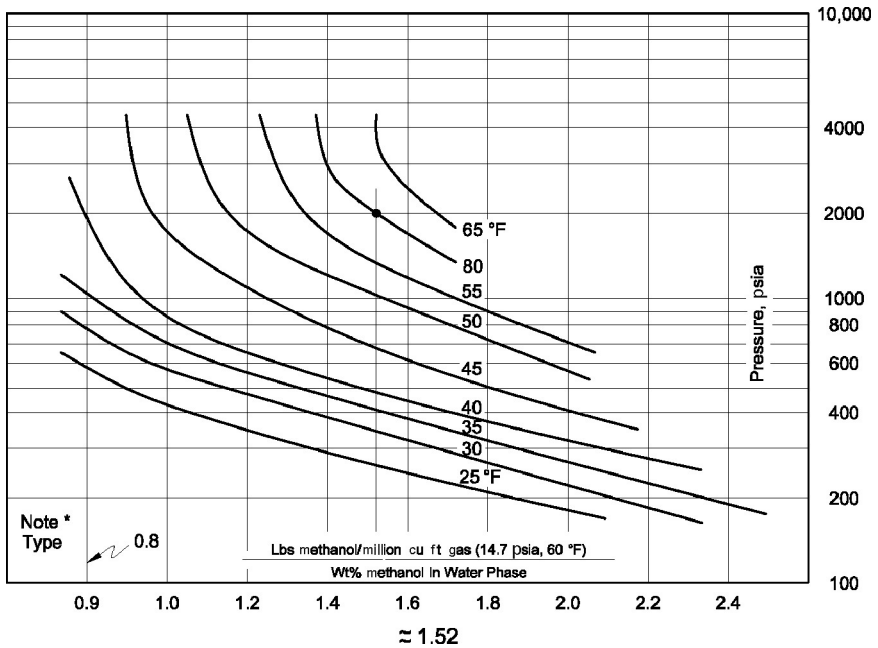


Figure 5.24 Ratio of methanol vapor composition to methanol liquid composition-example.

5. From Figure 5.24, the methanol that will flash into the vapor phase at 2000 psia and 60 °F is:

$$= \frac{(x) \text{ lbs. methanol} = \text{MMSCF} (@14.7 \text{ psia and } 60^\circ \text{F})}{WT \text{ and methanol in water phase}} = 1.52$$

6. Therefore, the methanol in the vapor phase (x) is: $(1.521)(9.892\%) = 4.94 \text{ lb/MMSCF}$.
7. A barrel of our condensate weighs: $(0.739)(5.6146 \text{ cf/bbl})(62.41 \text{ lb/cf}) = 258.9 \text{ lb/bbl}$.
8. Therefore, the approximate amount of methanol soluble in the condensate or liquid hydrocarbon phase (assuming a 0.5% solubility by weight) is:

$$= \frac{0.005}{1 - 0.005} (258.9 \text{ lb/bbl})(40 \text{ bbl/MMSCF}) = 52.04 \text{ lb/MMSCF}$$

9. Thus, the total amount of methanol required is:

$$\begin{aligned} \text{Liquid water phase} &= 121.15 \text{ lb/MMSCF} \\ \text{Vapor phase} &= 14.94 \text{ lb/MMSCF} \end{aligned}$$

$$\text{Soluble in condensate} = 52.04 \text{ lb/MMSCF}$$

$$\text{Total} = 188.13 \text{ lb/MMSCF}$$

$$\text{lb/day} = (188.13 \text{ lb/MMSCF}) \times (20 \text{ MMSCFD})$$

$$= 3762.6 \text{ lb/day}$$

Note that for gas-condensate wells producing a reasonable or high amount of condensate, the amount of methanol soluble in the condensate is crucial to determining the amount needed.

Approximately 188 lb of methanol must be added so that ~121 lb will be dissolved into the water phase. Since the specific gravity of methanol is 0.791 (at 68 °F), this is equivalent to:

$$\frac{188.13 \text{ lb/MMSCF}}{(0.791)(8.3453 \text{ lb/gal})} = 28.5 \text{ gal/MMSCF}$$

$$\frac{(28.5 \text{ gal/MMSCF})}{42 \text{ gal/bbl}} = 0.679 \text{ bbl/MMSCF}$$

$$(0.679 \text{ bbl/MMSCF})(20 \text{ MMSCF}) = 13.57 \text{ bbl/day}$$

Note the sensitivity of the total methanol requirements as the solubility of methanol is varied from 0.5 to 3.0 wt% in the following spreadsheet for our example (Figure 5.25). (Current research reports and laboratory analysis suggest the solubility is actually closer to 0.5 wt%.)



EXERCISES

1. Calculate the water content of the following gas stream at 2000 psia and 100 °F.

Component	Mol%
N ₂	8.5
H ₂ S	5.4
CO ₂	0.5
C ₁	77.6
C ₂	5.9
C ₃	1.9
iC ₄	0.1
nC ₄	0.1
iC ₅	100.0

Hydrate Inhibition-Hammerschmidt's Equation
Minimum Inhibitor Requirements-Methanol
Works Rev.3: GPSA (Fig. 20-33, p. 20-18)

Note the sensitivity of the overall MeOH volume based on amount of condensate and the solubility of the MeOH into the condensate liquid phase!

Input values:

CASE I.D.:	Well #1	Well #1	Well #1	Well #1	Well #1
Qg = Nominal Gas Flow Rate (MMSCFD)	20	20	20	20	20
Ql = Condensate Flow Rate (bbl/day)	800	800	800	800	800
Qw = Free Water Flow Rate (bbl/day)	60	60	60	60	60
SGg = Gas Specific Gravity	0.600	0.600	0.600	0.600	0.600
SGc = Condensate Specific Gravity	0.739	0.739	0.739	0.739	0.739
SGw = Free Water Specific Gravity	1.030	1.030	1.030	1.030	1.030
SGm = Methanol Specific Gravity	0.791	0.791	0.791	0.791	0.791
Wc% = MeOH Solubility In Condensate (wt%)	3.00	1.50	1.00	0.80	0.53
W1 = Sat. Water Content Upstream Of Choke (lb/MMSCF)	32.0	32.0	32.0	32.0	32.0
W2 = Sat. Water Content Downstream Of Choke (lb/MMSCF)	11.5	11.5	11.5	11.5	11.5
P2 = Pressure Downstream Of Choke (psig)	2000	2000	2000	2000	2000
T2 = Temperature Downstream Of Choke (°F)	60.0	60.0	60.0	60.0	60.0
dT = Desired Temperature Margin (°F) Above Hydrate Formation Temperature (i.e., Safety Factor)	0	0	0	0	0

Intermediate values:

Wf = Free Water (lb/day)	21,653	21,653	21,653	21,653	21,653
Wc = Water Condensed (lb/day)	410	410	410	410	410
Wo = Total Water Flow Rate	22,063	22,063	22,063	22,063	22,063
R = Methanol Vapor/Liquid Ratio (lbs MeOH/MMSCF/wt%)	1.508	1.508	1.508	1.508	1.508
d = Req'd. Water Dewpoint Depression (°F)	8.35	8.35	8.35	8.35	8.35
Kh = Inhibitor Constant (lb-°F/lbmol)	2335	2335	2335	2335	2335
MWl = Methanol Mol. Wgt. (lb/mol)	32.042	32.042	32.042	32.042	32.042
= Req'd. Weight % Inhibitor In Liquid Water Phase	10.28	10.28	10.28	10.28	10.28
Ww = MeOH Lost To Vapor Phase (lb/day)	310	310	310	310	310
Wl = MeOH In Water Liquid Phase (lb/day)	2527	2527	2527	2527	2527
Wc = MeOH In Condensate Liquid Phase (lb/day)	6406	3154	2092	1670	1111

Output values:

Th = Gas Hydrate Temperature (°F)	68.3	68.3	68.3	68.3	68.3
Wm = Req'd. Methanol Mass Flow Rate (lb/day)	9243	5991	4929	4507	3948
(lb/h)	385	250	205	188	164
(lb/MMSCF)	462	300	246	225	197
Qm = Req'd. Methanol Liquid Volume Flow Rate (gal/day)	1,401	908	747	683	598
(bbl/day)	33.3	21.6	17.8	16.3	14.2
(gal/Min)	0.97	0.63	0.52	0.47	0.42
(gal/MMSCF)	70.07	45.39	37.35	34.15	29.91
(bbl/MMSCF)	1.67	1.08	0.89	0.81	0.71

Figure 5.25 Spreadsheet illustrating the sensitivity of the solubility of methanol.

Use the following:

- (a) McKetta-Wehe graph
 - (b) Weighed average method (Equation 5.2)
 - (c) SRK “quick look” method
2. A gas stream, saturated with water, leaves the wellhead at 122 °F and 2900 psia. Some distance downstream, the gas enters a separator at 1015 psia and 50 °F. How much liquid water should be drained from the separator, if any?

3. Given a gas stream at 1000 psia, $MW=20.37$ and the following composition:

Component	Mol%
N ₂	10.1
C ₁	77.1
C ₂	6.1
C ₃	3.5
<i>i</i> C ₄	0.7
<i>n</i> C ₄	1.1
C ₅ +	0.8
	100.0

Determine the hydrate formation temperature using:

- (a) Vapor-solid equilibrium constant
 - (b) Pressure-temperature correlation curves
4. Determine the temperature drop across a choke with a flowing tubing pressure of 5000 psia and a downstream back pressure of 1000 psia. The well stream produces 60 bbl/MMSCFD of liquid hydrocarbon.
5. 9.5 MMSCFD of natural gas ($S=0.65$) having a hydrate formation temperature of 70 °F cools to 40 °F in a burned pipeline. Assume pipeline pressure is 900 psia. How much methanol must be added in bbl/day, if the gas enters the line saturated at 90 °F and is free of liquid water.

Condensate Stabilization

The liquids that are separated from the gas stream in the first separator may be flowed directly to a tank or may be “stabilized” in some fashion. As was discussed in Chapter 2 of Volume 1, these liquids contain a large percentage of methane and ethane, which will flash to gas in the tank. This lowers the partial pressure of all other components in the tank and increases their tendency to flash to vapors. The process of increasing the amount of intermediate (C_3 - C_5) and heavy (C_3 - C_5) and heavy (C_6^+) components in the liquid phase is called “stabilization.” In a gas field this process is called *condensate stabilization*, and in an oil field it is called *crude stabilization*.

In almost all cases, the molecules have a higher value as liquid than as gas. Crude oil streams typically contain a low percentage of intermediate components. Thus, it is not normally economically attractive to consider other alternatives to multistage separation to stabilize the crude. In addition, the requirement to treat the oil at a high temperature is more important than stabilizing the liquid and may require the flashing of both intermediate and heavy components to the gas stream.

Gas condensate, on the other hand, may contain a relatively high percentage of intermediate components and can be easily separated from entrained water due to its lower viscosity and greater density difference with water. Thus, some sort of condensate stabilization should be considered for each gas well production facility.



6.1 PARTIAL PRESSURES

As pointed out in Volume 1, the fraction of any one component that flashes to gas at any stage in a process is a function of the temperature, pressure, and composition of the fluid at that stage. For a given temperature, this tendency to flash can be visualized by the partial pressure of the component in the gas phase that is in equilibrium with the liquid. Partial pressure is defined as:

$$\text{Partial Pressure}_n = \frac{\text{Moles}_n}{\text{Sum of moles}} \times \text{Gas Pressure}$$

The partial pressure at a given pressure and temperature is lower when there are more moles of other components in the gas phase. The lower the partial pressure, the greater the tendency of the component to flash to gas. Thus, the higher the fraction of light components in the inlet fluid to any separator, the lower the partial pressure of intermediate components in the gas phase of the separator, and the greater the number of intermediate component molecules that flash to gas.



6.2 MULTISTAGE SEPARATION

Figure 6.1 shows a multistage separation process. By removing molecules of the light components in the first separator, these components are not available to flash to gas from the liquid in the second separator, and the partial pressure of intermediate components in the second separator is higher than it would have been if the first separator did not exist. The second separator serves the same function of increasing the partial pressure of the intermediate components in the third separator, and so forth.

The simplest form of condensate stabilization is to install a low-pressure separator downstream of an initial high-pressure separator. Unless the gas well produces at low pressure (<500 psi) and the gas contains very little condensate (<100 bpd), the additional expenditure for this stage of separation is almost always economical when balanced against increased liquid

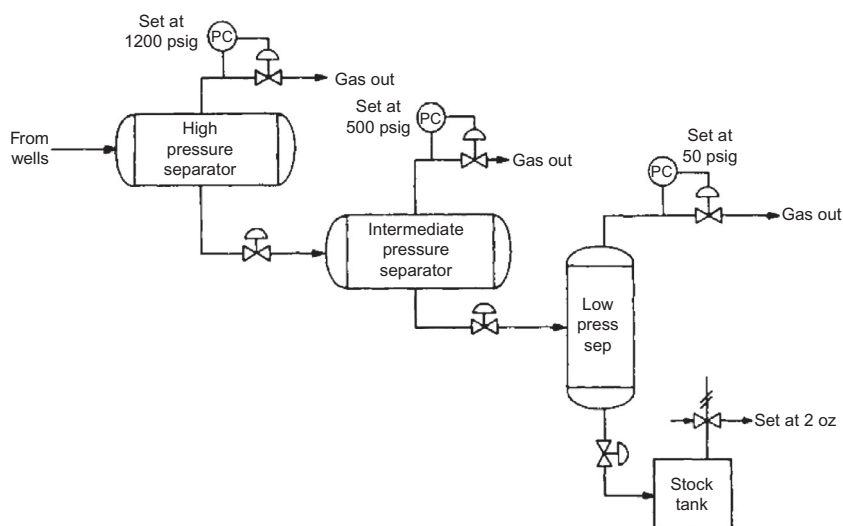


Figure 6.1 Multistage separation process.

production. If vapor recovery from the tank is required by environmental regulations, the flash separator will significantly reduce the horsepower required. If vapor recovery is not required, the gas from the flash separator may be economically feasible to be recovered and recompressed for sales even if it is not feasible to recover stock-tank vapors.

6.3 MULTIPLE FLASHES AT CONSTANT PRESSURE AND INCREASING TEMPERATURE

It is possible to stabilize a liquid at a constant pressure by successively flashing it at increasing temperatures as shown in Figure 6.2. At each successive stage, the partial pressure of the intermediate components is higher than it could have been at that temperature if some of the lighter components had not been removed by the previous stage. It would be very costly to arrange a process as shown in Figure 6.2, and this is never done. Instead, the same effect is obtained in a tall, vertical pressure vessel with a cold temperature at the top and a hot temperature at the bottom. This is called a “condensate stabilizer.”

Figure 6.3 shows a condensate stabilizer system. The well stream flows to a high pressure, three-phase separator. Liquids containing a high fraction of light ends are cooled and enter the stabilizer tower at ~ 200 psi.

In the tower the liquid falls downward in a process that results in many flashes at ever-increasing temperatures. At the bottom of the tower, some of

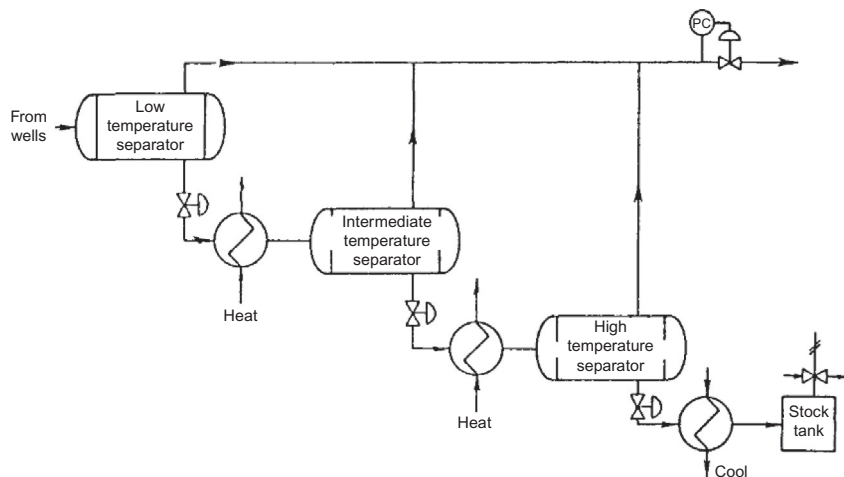


Figure 6.2 Multiple flashes at constant pressure and increasing temperature.

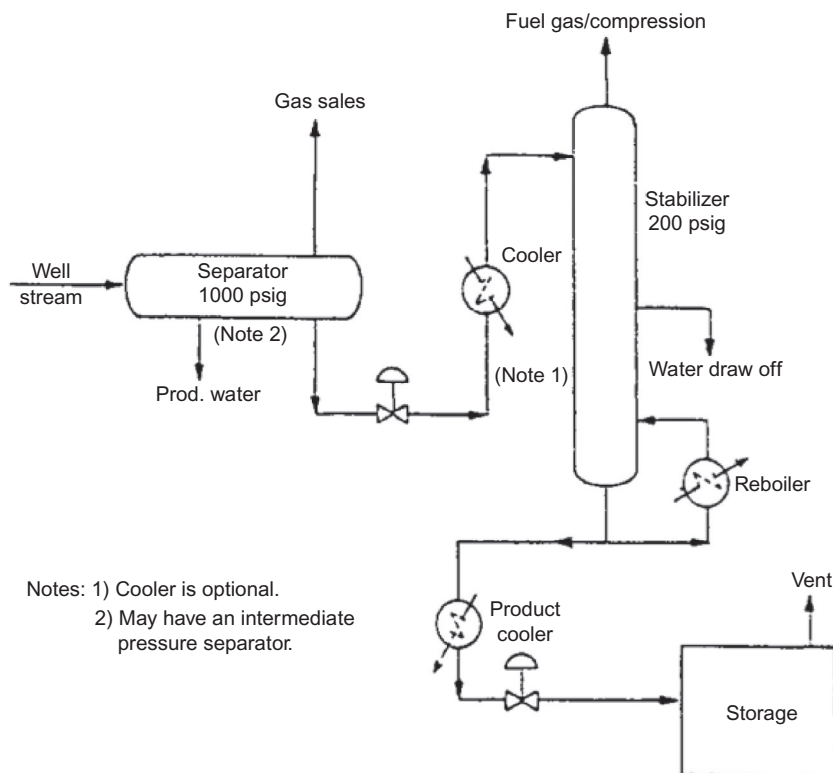


Figure 6.3 Condensate stabilization system.

the liquids are cycled to a reboiler where they receive heat to provide the necessary bottoms temperature (200–400 °F). The reboiler could be either a direct-fired bath, an indirect-fired bath, or a heat medium exchanger.

The liquids leaving the bottom of the tower have undergone a series of stage flashes at ever-increasing temperatures, driving off the light components, which exit the top of the tower. These liquids must be cooled to a sufficiently low temperature to keep vapors from flashing to atmosphere in the storage tank.



6.4 COLD-FEED DISTILLATION TOWER

Figure 6.4 shows the cold-feed distillation tower of Figure 6.3. The inlet stream enters the top of the tower. It is heated by the hot gases bubbling up through it as it falls from tray to tray through the downcomers. A flash

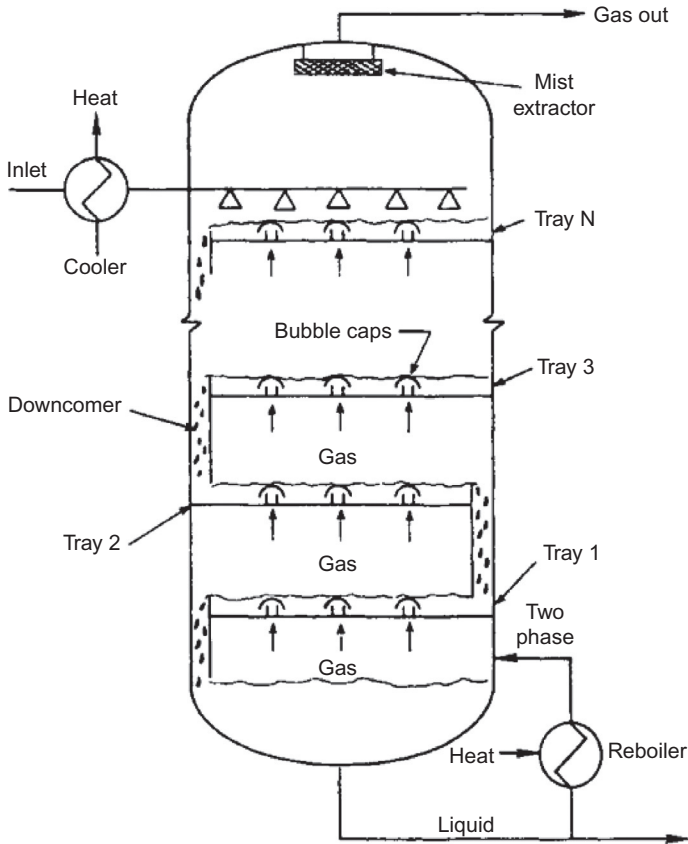


Figure 6.4 Cold-feed distillation tower of condensate stabilization system.

occurs on each tray so that the liquid is in near equilibrium with the gas above it at the tower pressure and the temperature of that particular tray.

As the liquid falls, it becomes leaner and leaner in light ends, and richer and richer in heavy ends. At the bottom of the tower, some of the liquid is circulated through a reboiler to add heat to the tower. As the gas goes up from tray to tray, more and more of the heavy ends get stripped out of the gas at each tray, and the gas becomes richer and richer in the light ends and leaner and leaner in the heavy ends (just the opposite of the liquid). The gas exits the top of the tower.

The lower the temperature of the inlet liquid, the lower the fraction of intermediate components that flash to vapor on the top trays and the greater the recovery of these components in the liquid bottoms. However, the colder the feed, the more heat is required from the reboiler to remove light

components from the liquid bottoms. If too many light components remain in the liquid, the vapor pressure limitations for the liquid may be exceeded. Light components may also encourage flashing of intermediate components (by lowering their partial pressure) in the storage tank. There is a balance between the amount of inlet cooling and the amount of reboiling required.

Typically, the liquid at the bottom of the tower must meet a specified vapor pressure. The tower must be designed to maximize the molecules of intermediate components in the liquid without exceeding the vapor pressure specification. This is accomplished by driving the maximum number of molecules of methane and ethane out of the liquid and keeping as much of the heavier ends as possible from going out with the gas.

Given inlet composition, pressure, and temperature, a tower temperature and the number of trays that produce a liquid with a specified vapor pressure can be chosen as follows:

1. Assume an initial split of components in the inlet that yields the desired vapor pressure. That is, assume a split of each component between the tower overhead (gas) and bottoms (liquid). There are various rules of thumb that can be used to estimate this split in order to give a desired vapor pressure. Once the split is made, both the assumed composition of the liquid and the assumed composition of the gas are known.
2. Calculate the temperature required at the base of the tower to develop this liquid. This is the temperature at the bubble point for the tower pressure and for the assumed outlet composition. Because the composition and pressure are known, the temperature at its bubble point can be calculated.
3. Calculate the composition of the gas in equilibrium with the liquid. The composition, pressure, and temperature of the liquid are known, and the composition of the gas that is in equilibrium with this liquid can be calculated.
4. Calculate the composition of the inlet liquid falling from Tray 1. Since the composition of the bottom liquid and gas in equilibrium with the liquid is known, the composition of the feed to this tray is also known. This is the composition of the liquid falling from Tray 1.
5. Calculate the temperature of Tray 1. From an enthalpy balance, the temperature of the liquid falling from Tray 1, and thus the temperature of the flash on Tray 1, can be calculated. If the composition is known, the enthalpy can be calculated. Enthalpy must be maintained, so the enthalpy of the liquid of known composition falling from Tray 1 must equal the sum of the enthalpies of the liquid and gas flashing from it at known temperatures.

6. This procedure can then be carried on up the tower to Tray N , which establishes the temperature of the inlet and the gas outlet composition.
7. From the composition of the inlet and gas outlet, the liquid outlet composition can be calculated and compared to that assumed in step 1.
8. The temperature or number of trays can then be varied until the calculated outlet liquid composition equals the assumed composition, *and* the vapor pressure of the liquid is equal to or less than that assumed. If the vapor pressure of the liquid is too high, the bottoms temperature must be increased.

6.5 DISTILLATION TOWER WITH REFLUX

Figure 6.5 shows a stabilizer with reflux. The well fluid is heated with the bottoms product and injected into the tower, below the top, where the temperature in the tower is equal to the temperature of the feed. This minimizes the amount of flashing. In the tower, the action is the same as in a cold-feed stabilizer or in any other distillation tower. As the liquid falls, the composition of the bottom liquid and gas in equilibrium through the tower goes from tray to tray and gets increasingly richer in the heavy components and increasingly leaner in the light components. The stabilized liquid is cooled in the heat exchanger by the feed stream before flowing to the stock tank.

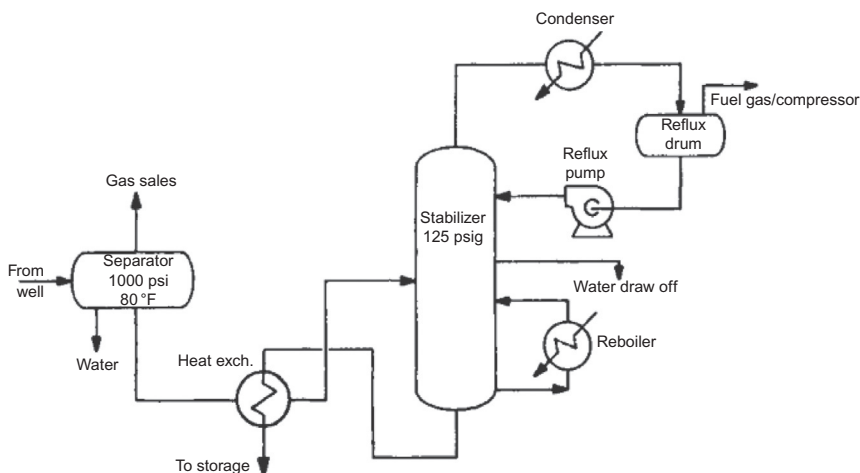


Figure 6.5 Stabilizer with reflux and feed/bottoms heat exchanger.

At the top of the tower any intermediate components going out with the gas are condensed, separated, pumped back to the tower, and sprayed down on the top tray. This liquid is called “reflux,” and the two-phase separator that separates it from the gas is called a “reflux tank” or “reflux drum.” The reflux performs the same function as the cold feed in a cold-feed stabilizer. Cold liquids strip out the intermediate components from the gas as the gas rises.

The heat required at the reboiler depends on the amount of cooling done in the condenser. The colder the condenser, the purer the product and the larger the percentage of the intermediate components that will be recovered in the separator and kept from going out with the gas. The hotter the bottoms, the greater the percentage of light components will be boiled out of the bottoms liquid and the lower the vapor pressure of the bottoms liquid.

A condensate stabilizer with reflux will recover more intermediate components from the gas than a cold-feed stabilizer. However, it requires more equipment to purchase, install, and operate. This additional cost must be justified by the net benefit of the incremental liquid recovery, less the cost of natural gas shrinkage and loss of heating value, over that obtained from a cold-feed stabilizer.



6.6 CONDENSATE STABILIZER DESIGN

It can be seen from the previous description that the design of both a cold-feed stabilizer and a stabilizer with reflux is a rather complex and involved procedure. Distillation computer simulations are available that can be used to optimize the design of any stabilizer if the properties of the feed stream and desired vapor pressure of the bottoms product are known. Cases should be run of both a cold-feed stabilizer and one with reflux before a selection is made. Because of the large number of calculations required, it is not advisable to use hand calculation techniques to design a distillation process. There is too much opportunity for computational error.

Normally, the crude or condensate sales contract will specify a maximum Reid vapor pressure (RVP). This pressure is measured according to a specific ASTM testing procedure. A sample is placed in an evacuated container such that the ratio of the vapor volume to the liquid volume is 4 to 1. The sample is then immersed in a 100F liquid bath. The absolute pressure measured is the RVP of the mixture.

Because a portion of the liquid was vaporized to the vapor space, the liquid will have lost some of its lighter components. This effectively changes

the composition of the liquid and yields a slightly lower vapor pressure than the true vapor pressure of the liquid at 100 °F. Figure 6.6 can be used to estimate true vapor pressure at any temperature from a known RVP.

The inherent error between true vapor pressure and RVP means that a stabilizer designed to produce a bottoms liquid with a true vapor pressure

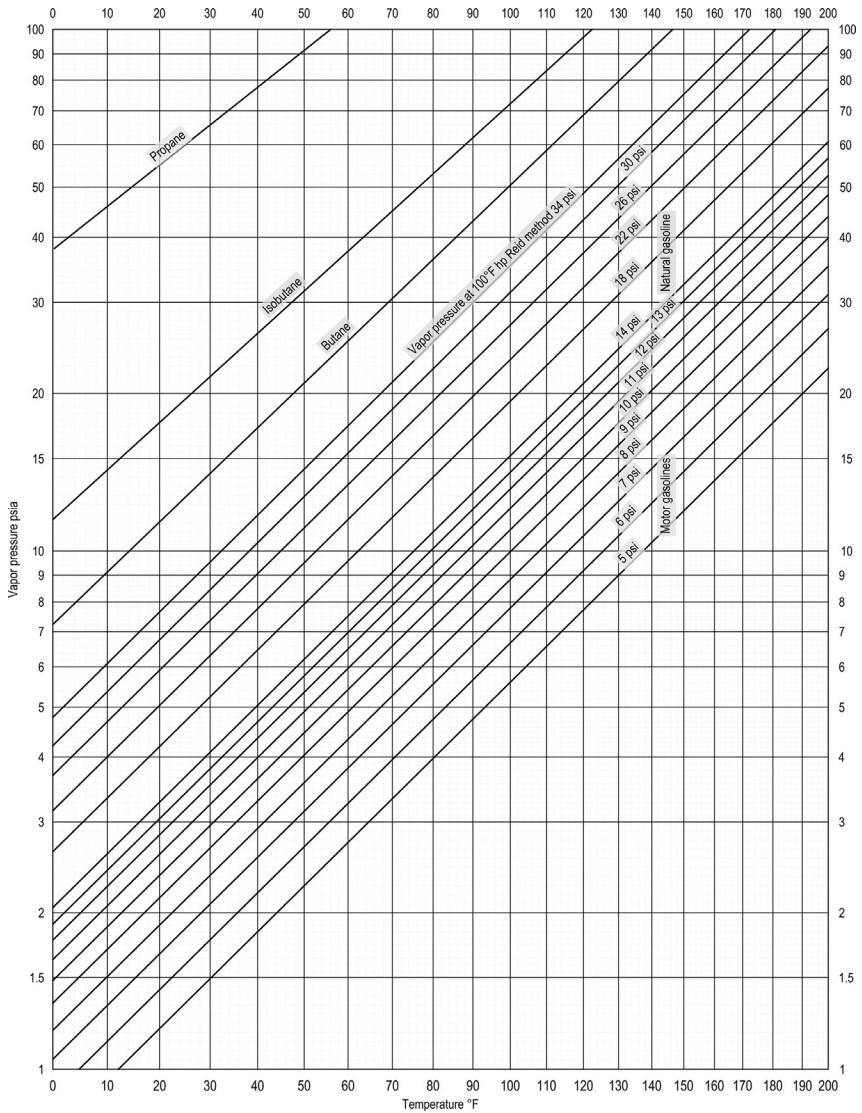


Figure 6.6 Relationship between Reid vapor pressure and actual vapor pressure. From Gas Processors Suppliers Association, *Engineering Data Book*, 9th Edition.

Table 6.1 Vapor Pressure and Relative Volatility of Various Components

Component	Vapor Pressure at 100 °F psia	Relative Volatility
C ₁	5000	96.9
C ₂	800	15.5
C ₃	190	3.68
<i>i</i> -C ₄	72.2	1.40
<i>n</i> -C ₄	51.6	1.00
<i>i</i> -C ₅	20.4	0.40
<i>n</i> -C ₅	15.6	0.30
C ₆	5.0	0.10
C ₇ ⁺	≈0.1	0.00
CO ₂	-	Infinite
N ₂	-	Infinite
H ₂ S	394	7.64

equal to the specified RVP will be conservatively designed. The vapor pressures of various hydrocarbon components at 100 °F are given in Table 6.1.

The bottoms temperature of the tower can be approximated if the desired vapor pressure of the liquid is known. The vapor pressure of a mixture is given by:

$$VP = \text{Sum}[VP_n \times MF_n] \quad (6.1)$$

where VP = vapor pressure of mixture, psia; VP_{*n*} = vapor pressure of component *n*, psia; MF_{*n*} = mole fraction of component *n* in liquid.

To estimate the desired composition of the bottom liquid, the vapor pressures of the different components at 100 °F can be assumed to be a measure of the volatility of the component. Thus, if a split of *n*-C₄ is assumed, the mole fraction of each component in the liquid can be estimated from:

$$L_n = \frac{F_n(n - C_4 \text{ split})}{RV_n} \quad (6.2)$$

$$MF_n = \frac{L_n}{\text{SUM}(L_n)} \quad (6.3)$$

where *F_n* = total number of moles of component *n* in feed; *L_n* = total number of moles of component *n* in the bottom liquid; (*n*-C₄ split) = assumed moles of component *n*-C₄ in bottom liquid divided by moles of *n*-C₄ in feed; RV_{*n*} = relative volatility of component *n* from Table 6.1.

To determine the composition of the bottom liquid, assume a split of *n*-C₄ and compute MF_{*n*} from Equations (6.2) and (6.3). The vapor pressure

can then be computed from Equation (6.1). If the vapor pressure is higher than the desired RVP, choose a lower number for the n -C₄ split. If the calculated vapor pressure is lower than the desired RVP, choose a higher number for the n -C₄ split. Iterate until the calculated vapor pressure equals the desired RVP.

The bottoms temperature can then be determined by calculating the bubble point of the liquid described by the previous iteration at the chosen operating pressure in the tower. This is done by choosing a temperature, determining equilibrium constants from Chapter 3, Volume 1, and computing:

$$C = \text{Sum} (L_n \times K_n) \quad (6.4)$$

If C is >1.0 , the assumed temperature is too high. If C is lower than 1.0, the assumed temperature is too low. By iteration a temperature can be determined where $C=1.0$.

Typically, bottoms temperatures will range from 200 to 400 °F depending on operating pressure, bottoms composition, and vapor pressure requirements. Temperatures should be kept to a minimum to decrease the heat requirements, limit salt buildup, and prevent corrosion problems.

When stabilizer operating pressures are kept below 200 psig, the reboiler temperatures will normally be below 300 °F. A water glycol heating medium can then be used to provide the heat. Higher stabilizer operating pressures require the use of steam- or hydrocarbon-based heating mediums. Operating at higher pressures, however, decreases the flashing of the feed on entering the column, which decreases the amount of feed cooling required. In general, a crude stabilizer should be designed to operate between 100 and 200 psig.



6.7 TRAYS AND PACKING

The number of actual equilibrium stages determines the number of flashes that will occur. The more stages, the more complete the split, but the taller and more costly the tower. Most condensate stabilizers will contain approximately five theoretical stages. In a refluxed tower, the section above the feed is known as the rectification section, while the section below the feed is known as the stripping section. The rectification section normally contains about two equilibrium stages above the feed, and the stripping section normally contains three equilibrium stages.

Theoretical stages within a tower are provided by actual stage devices (typically either trays or packings). The actual diameter and height of the tower can be derived using the manufacturer's data for the particular device. The height of the tower is a function of the number of theoretical stages and of the efficiency of the actual stages. The diameter of the tower is a function of the hydraulic capacity of the actual stages.

6.7.1 Trays

For most trays, liquid flows across an “active area” of the tray and then into a “downcomer” to the next tray below, and so on. Inlet and/or outlet weirs control the liquid distribution across the tray. Vapor flows up the tower and passes through the tray active area, bubbling up through (and thus contacting) the liquid flowing across the tray. The vapor distribution is controlled by (1) perforations in the tray deck (sieve trays), (2) bubble caps (bubble cap trays), or (3) valves (valve trays).

Trays operate within a hydraulic envelope. At excessively high vapor rates, liquid is carried upward from one tray to the next (essentially backmixing the liquid phase in the tower). For valve trays and sieve trays, a capacity limit can be reached at low vapor rates when liquid falls through the tray floor rather than being forced across the active area into the downcomers. Because the liquid does not flow across the trays, it misses contact with the vapor, and the separation efficiency drops dramatically.

Trays are generally divided into four categories: (1) sieve trays, (2) valve trays, (3) bubble cap trays, and (4) high-capacity/high-efficiency trays.

6.7.1.1 Sieve Trays

Sieve trays are the least expensive tray option. In sieve trays, vapor flowing up through the tower contacts the liquid by passing through small perforations in the tray floor (Figure 6.7b). Sieve trays rely on vapor velocity to exclude liquid from falling through the perforations in the tray floor. If the vapor velocity is much lower than design, liquid will begin to flow through the perforations rather than into the downcomer. This condition is known as weeping. Where weeping is severe, the equilibrium efficiency will be very low. For this reason, sieve trays have a very small turndown ratio.

6.7.1.2 Valve Trays

Valve trays are essentially modified sieve trays. Like sieve trays, holes are punched in the tray floor. However, these holes are much larger than those

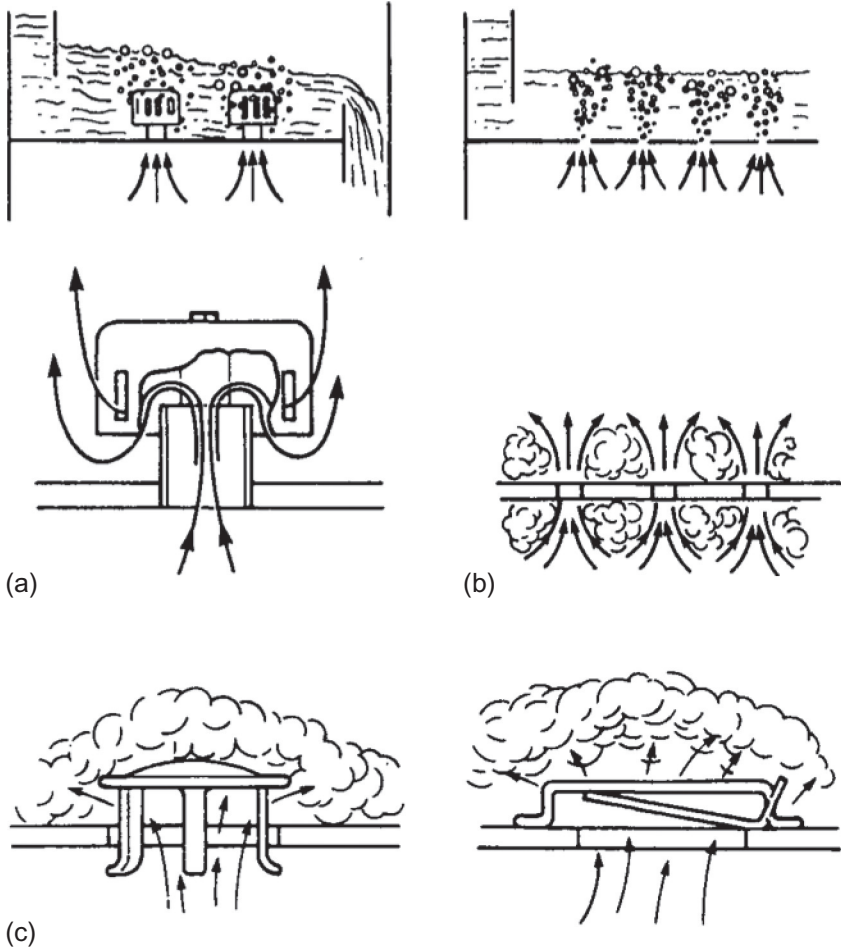


Figure 6.7 Vapor flow through trays. Vapor flow through (a) bubble cap, (b) perforations, and (c) valves.

in sieve trays. Each of these holes is fitted with a device called a “valve.” Vapor flowing up through the tower contacts the liquid by passing through valves in the tray floor (Figure 6.7c). Valves can be fixed or moving. Fixed valves are permanently open and operate as deflector plates for the vapor coming up through the holes in the tray floor. For moving valves, vapor passing through the tray floor lifts the valves and contacts the liquid. Moving valves come in a variety of designs, depending on the manufacturer and the application. At low vapor rates, valves will close, helping to keep liquid from falling through the holes in the deck. At sufficiently low vapor rates, a valve

tray will begin to weep. That is, some liquid will leak through the valves rather than flowing to the tray downcomers. At very low vapor rates, it is possible that all the liquid will fall through the valves and no liquid will reach the downcomers. This severe weeping is known as “dumping.” At this point, the efficiency of the tray is nearly zero.

6.7.1.3 Bubble Cap Trays

In bubble cap trays, vapor flowing up through the tower contacts the liquid by passing through bubble caps (Figure 6.7a). Each bubble cap assembly consists of a riser and a cap. The vapor rising through the column passes up through the riser in the tray floor and then is turned downward to bubble into the liquid surrounding the cap. Because of their design, bubble cap trays cannot weep. However, bubble cap trays are also more expensive and have a lower capacity/higher pressure drop than valve trays or sieve trays.

6.7.1.4 High-Capacity/High-Efficiency Trays

High-capacity/high-efficiency trays have valves or sieve holes or both. They typically achieve higher efficiencies and capacities by taking advantage of the active area under the downcomer. At this time, each of the major vendors of high-capacity/high-efficiency trays has its own version of these trays, and the designs are proprietary.

6.7.1.5 Bubble Cap Trays Versus Valve Trays

At low vapor rates, valve trays will weep. Bubble cap trays cannot weep (unless they are damaged). For this reason, it is generally assumed that bubble cap trays have nearly an infinite turndown ratio. This is true in absorption processes (e.g., glycol dehydration), in which it is more important to contact the vapor with liquid than the liquid with vapor. However, this is not true of distillation processes (e.g., stabilization), in which it is more important to contact the liquid with the vapor.

As vapor rates decrease, the tray activity also decreases. Eventually, there comes a point at which some of the active devices (valves or bubble caps) become inactive. Liquid passing these inactive devices gets very little contact with vapor. At very low vapor rates, the vapor activity will concentrate only in certain sections of the tray (or, in the limit, one bubble cap or one valve). At this point, it is possible that liquid may flow across the entire active area without ever contacting a significant amount of vapor. This will result in very low tray efficiencies for a distillation process. Nothing can be done with a bubble cap tray to compensate for this.

However, a valve tray can be designed with heavy valves and light valves. At high vapor rates, all the valves will be open. As the vapor rate decreases, the valves will begin to close. With light and heavy valves on the tray, the heavy valves will close first, and some or all of the light valves will remain open. If the light valves are properly distributed over the active area, even though the tray activity is diminished at low vapor rates, what activity remains will be distributed across the tray. All liquid flowing across the tray will contact some vapor, and mass transfer will continue. Of course, even with weighted valves, if the vapor rate is reduced enough, the tray will weep and eventually become inoperable. However, with a properly designed valve tray this point may be reached after the loss in efficiency of a comparable bubble cap tray. So, in distillation applications, valve trays can have a greater vapor turndown ratio than bubble cap trays.

6.7.1.6 Tray Efficiency and Tower Height

In condensate stabilizers, trays generally have 70% equilibrium stage efficiency. That is, 1.4 actual trays are required to provide one theoretical stage. The spacing between trays is a function of the spray height and the downcomer backup (the height of clear liquid established in the downcomer). The tray spacing will typically range from 20 to 30 in. (with 24 in. being the most common), depending on the specific design and the internal vapor and liquid traffic. The tray spacing may increase at higher operating pressures (>165 psia) because of the difficulty in disengaging vapor from liquid on both the active areas and in the downcomers.

6.7.2 Packing

Packing typically comes in two types: random and structured.

Liquid distribution in a packed bed is a function of the internal vapor/liquid traffic, the type of packing employed, and the quality of the liquid distributors mounted above the packed bed. Vapor distribution is controlled by the internal vapor/liquid traffic, by the type of packing employed, and by the quality of the vapor distributors located below the packed beds.

Packing material can be plastic, metal, or ceramic. Packing efficiencies can be expressed as HETP (height equivalent to a theoretical plate).

6.7.2.1 Random Packing

A bed of random packing typically consists of a bed support (typically a gas injection support plate) upon which pieces of packing material are randomly arranged (they are usually poured or dumped onto this support plate). Bed

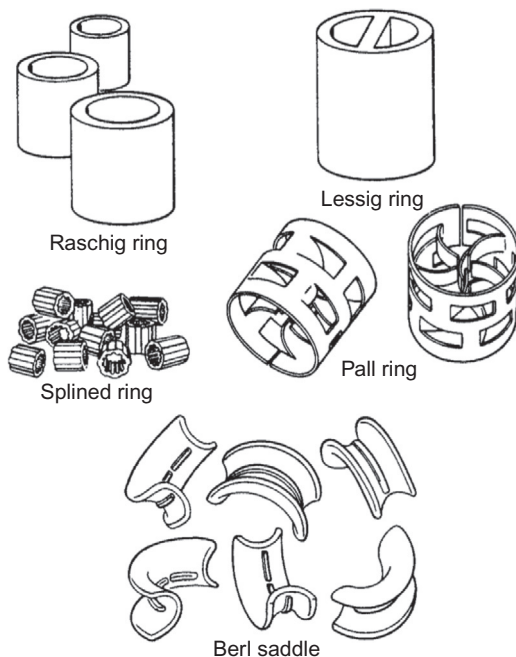


Figure 6.8 Various types of packing.

limiters, or hold-downs, are sometimes set above random beds to prevent the pieces of packing from migrating or entraining upward. Random packing comes in a variety of shapes and sizes. For a given shape (design) of packing, small sizes have higher efficiencies and lower capacities than large sizes.

Figure 6.8 shows a variety of random packing designs. An early design is known as a Raschig ring. Raschig rings are short sections of tubing and are low-capacity, low-efficiency, high-pressure drop devices. Today's industry standard is the slotted metal (Pall) ring. A packed bed made of 1-inch slotted metal rings will have a higher mass transfer efficiency and a higher capacity than will a bed of 1-inch Raschig rings. The HETP for a 2-inch slotted metal ring in a condensate stabilizer is about 36 inches. This is slightly more than a typical tray design, which would require 34 inches. ($1.4 \text{ trays} \times 24\text{-in. tray spacing}$) for one theoretical plate or stage.

6.7.2.2 Structured Packing

A bed of structured packing consists of a bed support upon which elements of structured packing are placed. Beds of structured packing typically have lower pressure drops than beds of random packing of comparable mass transfer efficiency. Structured packing elements are composed of grids (metal or plastic) or woven mesh (metal or plastic) or of thin vertical crimped sheets



Figure 6.9 Structured packing can offer better mass transfer than trays. *Courtesy of Koch Engineering Co., Inc.*

(metal, plastic, or ceramic) stacked parallel to each other. [Figure 6.9](#) shows examples of the vertical crimped sheet style of structured packing.

The grid types of structured packing have very high capacities and very low efficiencies, and are typically used for heat transfer or for vapor scrubbing. The wire mesh and the crimped sheet types of structured packing typically have lower capacities and higher efficiencies than the grid type.

6.7.3 Trays or Packing

There is no one answer for determining the type of tray or packing design. The choice is dictated by project scope (new tower or retrofit), current economics, operating pressures, anticipated operating flexibility, and physical properties.

6.7.3.1 Distillation Service

For distillation services, as in condensate stabilization, tray design is well understood, and many engineers are more comfortable with trays than with packing. In the past, bubble cap trays were the standard. However, they are no longer commonly used in this service. Sieve trays are inexpensive but offer a very narrow operating range when compared with valve trays. Although valve trays offer a wider operating range than do sieve trays, they have moving parts and so may require more maintenance. High-capacity/high-efficiency trays can be more expensive than standard valve trays. However, high-capacity/high-efficiency trays require smaller diameter towers, so they can offer significant savings in the overall cost of the distillation tower. The high-capacity/high-efficiency tray can also be an ideal choice for tower retrofits in which increased throughputs are required for existing towers.

Random packing has traditionally been used in small diameter (<20 in.) towers. That is because it is easier and less expensive to pack these small diameter towers. However, random packed beds are prone to channeling and have poor turndown characteristics when compared with trays. For these reasons, trays were preferred for tower diameters >20 in. In recent years, an improved understanding of the impact of high pressure on packing performance has been gained. Improved vapor and liquid distributor designs and modified bed heights have made the application of packing to large-diameter, high-pressure distillation towers more common. A properly designed packed bed system (packing, liquid distributors, vapor distributors) can be an excellent choice for debottlenecking existing distillation towers.

6.7.3.2 Stripping Service

For stripping service, as in a glycol or amine contactor (see [Chapters 7 and 8](#)), bubble cap trays are the most common. In recent years, there has been a growing movement toward crimped sheet structured packing. Improved vapor and liquid distributor design in conjunction with structured packing can lead to smaller-diameter and shorter stripping towers than can be obtained with trays.



6.8 CONDENSATE STABILIZER AS A GAS-PROCESSING PLANT

A gas-processing plant, as described in [Chapter 9](#), is designed to recover ethane, propane, butane, and other natural gas liquids (NGLs) from the gas stream. A condensate stabilizer also recovers some portion of these

liquids. The colder the temperature of the gas leaving the overhead condenser in a reflux stabilizer—or the colder the feed stream in a cold-feed stabilizer, and the higher the pressure in the tower—the greater the recovery of these components as liquids. Indeed, any stabilization process that leads to recovery of more molecules in the final liquid product is removing those molecules from the gas stream. In this sense, a stabilizer may be considered as a simple form of a gas-processing plant.

It is difficult to determine the point at which a condensate stabilizer becomes a gas plant. Typically, if the liquid product is sold as a condensate, the device would be considered a condensate stabilizer. If the product is sold as a mixed NGL stream or is fractionated into its various components, the same process would be considered a gas plant. The least volatile NGL stream has an RVP between 10 and 14 and has sufficient light hydrocarbons such that 25% of the total volume is vaporized at 140 °F.



6.9 LOW TEMPERATURE EXTRACTOR (LTX) UNIT AS A CONDENSATE STABILIZER

It should be clear from the description of LTX units in [Chapter 5](#) that the lower pressure separator in an LTX unit is a simple form of cold-feed condensate stabilizer. In the cold, upper portion of the separator some of the intermediate hydrocarbon components condense. In the hot, lower portion some of the lighter components flash.

An LTX unit is not a very efficient stabilizer because the absence of trays or packing keeps the two phases from approaching equilibrium at the various temperatures that exist in the vessel. In addition, it is difficult to control the process. Typically, for a 100–200-psi operating pressure, a 300–400 °F bottoms temperature is required to completely stabilize the condensate. The heating coil in an LTX separator is more likely to be in the range of 125–175 °F, and thus complete stabilization will not occur even if the flash were capable of reaching equilibrium.

There may be some additional recovery from an LTX unit than would be realized from a straight two-stage flash separation process, but this increment is normally small and may not justify the increased equipment cost and operating complexity associated with an LTX unit.



Dehydration



7.1 OVERVIEW

The major reasons for removing water from natural gas are as follows:

- Natural gas, at the correct pressure and temperature, can combine with free water to form solid hydrates that can plug valves, fittings, instruments, and pipelines. If hydrate prevention methods are unsuitable and hydrates are likely to form, some water must be removed from the gas stream.
- Water vapor increases the volume of the gas, which decreases the heating value of the gas.
- As gas travels in a pipeline, it is cooled, and water can condense. Slug flow could form, which could cause erosion and/or corrosion.
- Sales gas contracts, gas plant inlet feed requirements, and/or pipeline specifications usually must meet a specified water content (e.g., sales gas 7 lb H₂O per MMSCF).

Dehydration is the process of removing water from the gas stream. Water removal from gas can be accomplished by several processes. The two most common methods are adsorption (solid desiccant) and absorption (liquid desiccant (glycol)). These methods utilize mass transfer of the water molecule into a liquid solvent (glycol) or a crystalline structure (dry desiccant). A third method, refrigeration (cooling the gas), employs cooling to condense the water molecule to the liquid phase with a subsequent injection of inhibitor to prevent hydrate formation. Refrigeration technology is discussed in [Chapter 10](#). One less common method of dehydration will also be described, a nonregenerable dehydrator (calcium chloride brine unit).



7.2 ADSORPTION

7.2.1 Process Overview

Adsorption is a physical phenomenon that occurs when molecules of a gas are brought into contact with a solid surface and some of them condense on

the surface. Dehydration of a gas (or a liquid hydrocarbon) with a dry desiccant is an adsorption process in which water molecules are preferentially held by the desiccant and removed from the gas stream. Adsorption involves a form of adhesion between the surface of the solid desiccant and the water vapor in the gas. Water forms a thin film that is held to the desiccant surface by forces of attraction, not by chemical reaction. The desiccant is a solid, granulated dehydrating medium with a large effective surface area (large number of small pores) per unit weight.

Typical desiccants might have as much as 4 million square feet of surface area per pound. Commonly used desiccants include:

- Alumina
- Silica gel
- Molecular sieves (mol sieves)

Many grades and qualities of each of these substances are commercially available. Figure 7.1 is an enlargement of a molecular sieve particle.

7.2.2 Principles of Adsorption

The achievement of equilibrium on a small surface displays the following pattern: Some passing molecules will condense on the surface (physical as

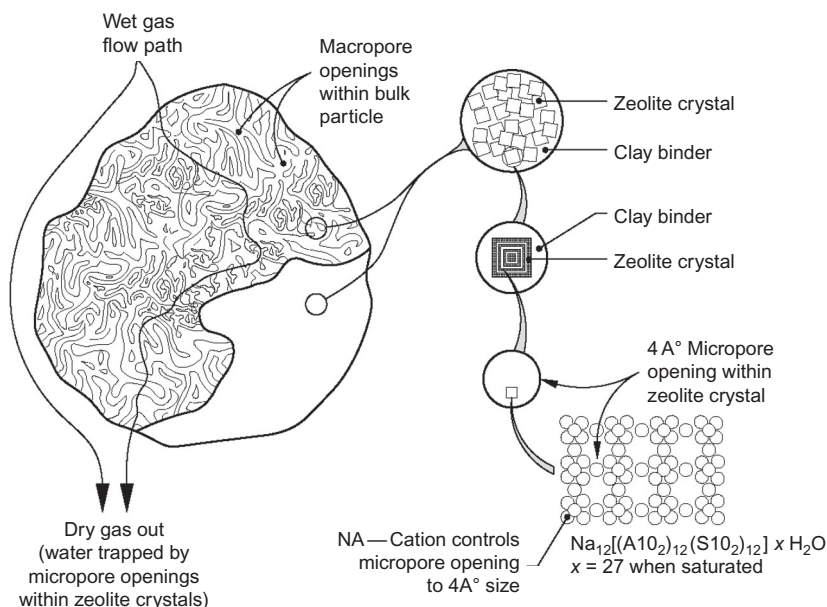


Figure 7.1 Enlargement of a molecular sieve particle.

opposed to chemical absorption). After some finite time, the molecule may acquire sufficient energy to leave and be replaced by another. After sufficient time, a state of equilibrium is reached wherein the number of molecules leaving the surface will equal the number arriving.

The number of molecules on the surface is a function of the nature of the adsorbent, the nature of the molecule being adsorbed (the adsorbate), the temperature of the system, and the concentration of the adsorbate over the adsorbent surface.

7.2.3 Process Reversal

The adsorption process may be reversed in the same manner that absorption processes are. Adsorption is encouraged by low temperatures and high pressures. Desorption (its reversal) is encouraged by high temperatures and low pressures.

7.2.4 Mass Transfer Zone

At the inlet of the bed and for a certain distance into it, the adsorbent is saturated to essentially equilibrium value with the adsorbable component in the fluid, such as water in natural gas. At the outlet of the bed, the adsorbent is unsaturated, and the water content of the gas is in equilibrium with the unsaturated activated adsorbent.

The mass transfer zone (MTZ) is defined as the zone between these two zones, where the concentration of the water in the natural gas is falling (Figure 7.2). MTZ lengths can be obtained experimentally for various materials and systems and used in graphical correlations for design purposes. MTZ is a function of the following factors: adsorbent, adsorbent particle size, fluid velocity, fluid properties, adsorbate concentration in the entering fluid, adsorbate concentration in the adsorbent if it is not fully reactivated, temperature, pressure, and past history of the system.

7.2.5 Principles of Operation

7.2.5.1 Introduction

The adsorption process is a batch process, with multiple desiccant beds used in cyclic operation to dry the gas on a continuous basis. The number and arrangement of the desiccant beds may vary from two towers, adsorbing alternatively (Figure 7.3), to many towers. Three separate functions, or cycles, must alternatively be performed in each dehydrator tower:

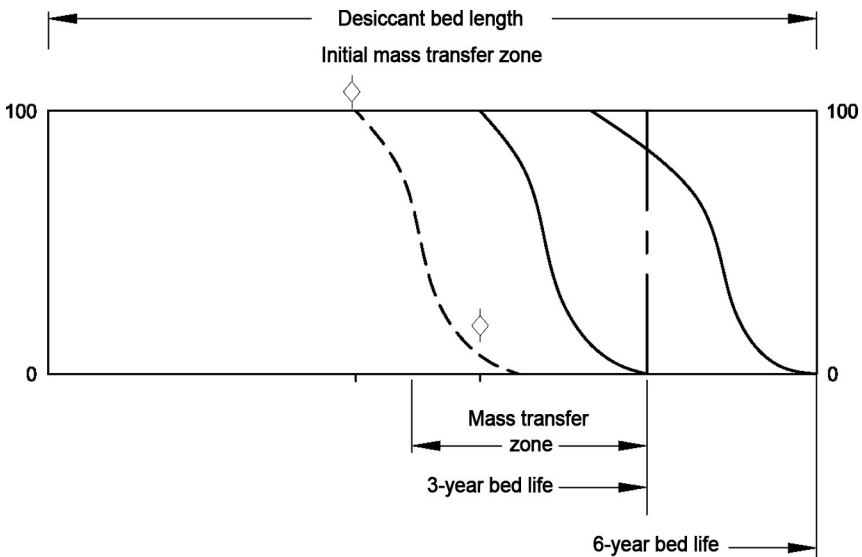


Figure 7.2 Schematic view of mass transfer zone (MTZ).

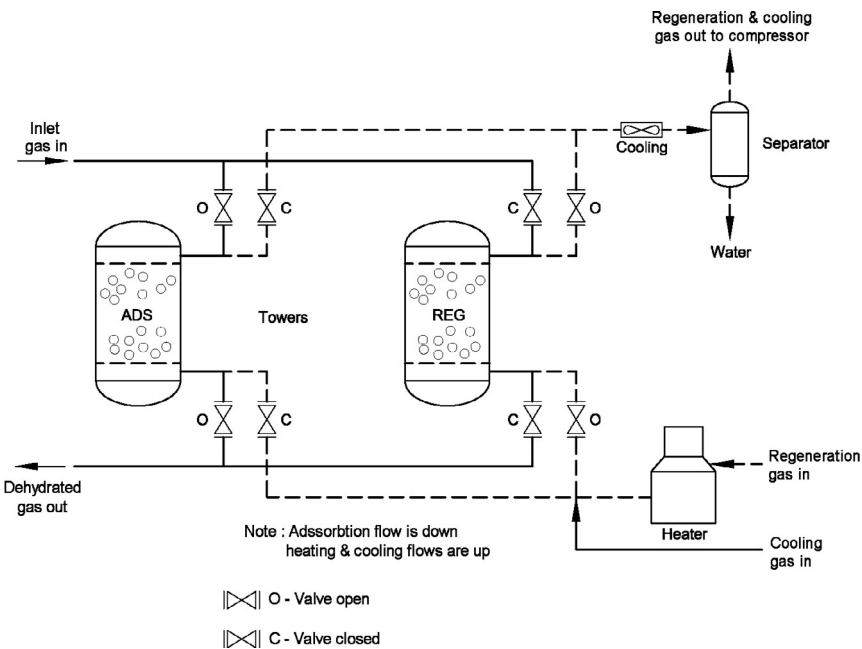


Figure 7.3 Simplified flow diagram of a two-tower solid desiccant dehydration system.

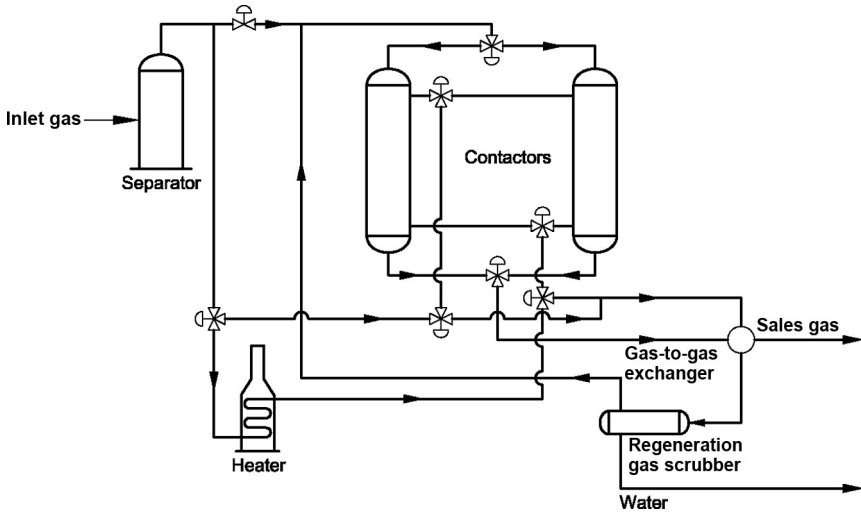


Figure 7.4 Flow diagram of a solid desiccant dehydration unit.

- Adsorbing or gas-drying cycle
- Heating or regeneration cycle
- Cooling cycle (prepares the regenerated bed for another adsorbing, or gas-drying, cycle)

Figure 7.4 is a flow diagram of a typical two-tower dehydration unit.

7.2.5.2 System Components

Essential components of a solid desiccant dehydration system are:

- Inlet gas stream microfiber filter separator
- Two or more adsorption towers (contactors) filled with a solid desiccant
- High-temperature heater to provide hot regeneration gas to reactivate the desiccant in the towers
- Regeneration gas cooler to condense water from the hot regeneration gas
- Regeneration gas separator to remove the condensed water from the regeneration gas
- Piping manifolds, switching valves, and controls to direct and control the flow of gases according to the process requirements

7.2.5.3 Drying/Reactivation Cycle

Figure 7.5 shows the flow of a typical two-tower unit with drying taking place in the first tower. Wet inlet gas first passes through an efficient

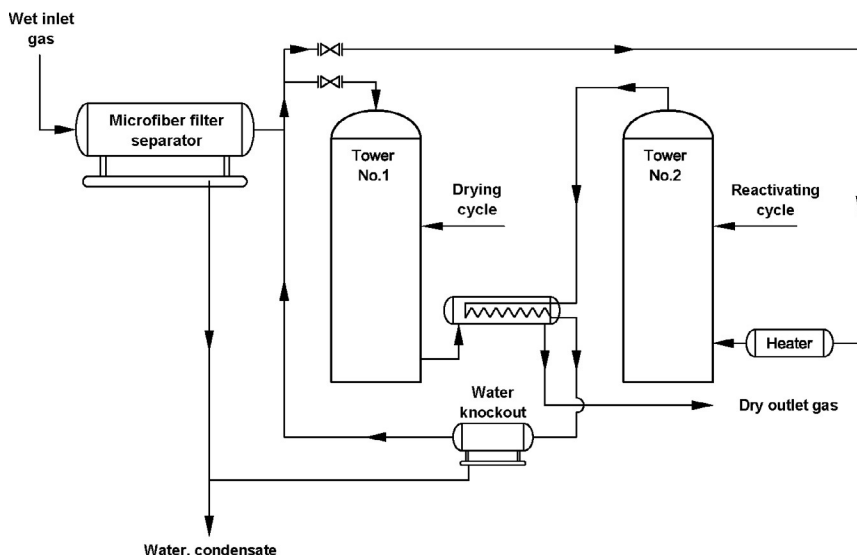


Figure 7.5 Flow diagram of a typical two-tower natural gas dehydrator.

microfiber inlet filter separator where free liquids, entrained mist, and solid particles are removed. Free liquids may damage or destroy the desiccant bed, and solids may plug the bed.

If the dehydration unit is downstream of an amine unit, glycol unit, or compressors, a microfiber filter inlet separator is highly recommended upstream of the adsorber towers. At any given time, one of the towers will be on stream in the adsorbing, or drying, cycle while the other is in the process of being heated or cooled. Several automatically operated switching valves and a controller route the inlet gas and regeneration gas to the correct tower at the proper time.

The tower being regenerated is heated for 5–6 h and cooled for the remaining 2–3 h.

As the wet gas flows downward through the tower on the adsorption cycle, each of the adsorbable components is adsorbed at a different rate. The water vapor is immediately adsorbed in the top layers of the desiccant bed. Some of the light hydrocarbon gases and heavier hydrocarbons moving down through the bed are also adsorbed. Heavier hydrocarbons will displace the lighter ones in the desiccant bed as the adsorbing cycle proceeds. As the upper layers of desiccant become saturated with water, water in the wet gas stream begins displacing the previously adsorbed hydrocarbons in the lower layers. For each component in the inlet gas stream, there will be a section of

bed depth, from top to bottom, where the desiccant is saturated with that component and where the desiccant below is just starting to adsorb it.

The depth of the bed from saturation to initial adsorption is the mass transfer zone. MTZ is simply a zone or section of the bed where a component is transferring its mass from the gas stream to the surface of the desiccant. As the flow of gas continues, the MTZs move downward through the bed, and water displaces all of the previously adsorbed gas until, finally, the entire bed is saturated with water vapor. When the bed is completely saturated with water vapor, the outlet gas is just as wet as the inlet gas.

Towers must be switched from the adsorption cycle to the regeneration cycle (heating and cooling) before the desiccant bed has become completely saturated. One regeneration-gas supply scheme consists of taking a portion (5–15%) of the entering wet gas stream across a pressure-reducing valve that forces a portion of the upstream gas through the regeneration system. In most plants, a flow controller regulates the volume of regeneration gas used. Regeneration gas is sent to a heater where it is heated to between 400 and 600 °F and then piped to the tower being regenerated. Initially, the hot regeneration gas must heat up the tower and the desiccant.

The water begins vaporizing when the effluent hot gas temperature reaches between 240 and 250 °F. The bed continues to heat up slowly as the water is being desorbed or driven out of the desiccant. After all the water has been removed, heating is maintained to drive off any heavier hydrocarbons and contaminants that would not vaporize at lower temperatures. The desiccant bed will be properly regenerated when the outlet gas (peak-out) temperature has reached between 350 and 550 °F. After the heating cycle, the desiccant bed is cooled by flowing unheated regeneration gas until the desiccant is sufficiently cooled.

All of the regeneration gas used in the heating and cooling cycles is passed through a heat exchanger (normally an aerial cooler) where it is cooled to condense the water removed from the regenerated desiccant bed. This water is separated in the regeneration gas separator, and the gas is mixed with the incoming wet gas stream. This entire procedure is continuous and automatic.

7.2.6 Performance

Advantages

- Can achieve very low dew points (<1 ppm)
- High contact temperatures are possible
- Adaptable to large rate and load changes

Disadvantages

- High initial cost
- Batch process
- Experiences high-pressure drop through the bed
- Desiccant is sensitive to poisoning with liquids or other impurities in the gas

Effect of Process Variables

Several process variables can have a major effect on dry bed dehydration sizing and operating efficiency. The variables include:

- Quality of inlet gas temperature pressure
- Cycle time
- Gas velocities
- Sources of regeneration gas
- Desiccant selection
- Effect of regeneration gas on outlet gas quality
- Pressure drop considerations

7.2.6.1 Quality of Inlet Gas

The performance of a dry bed dehydrator is affected by the moisture content of inlet gas and the components in the produced natural gas stream. The relative saturation of the inlet gas determines the size of a given desiccant bed and affects the transfer of water to the adsorbent.

Higher capacities can be expected when drying saturated gas (100% relative humidity) for most desiccants (except molecular sieve) than when drying partially saturated gases. In most gas field applications, the inlet gas is saturated with water vapor and thus this variable need not be considered.

Compounds in produced natural gas adversely affect the performance of the dry bed dehydrator. Components of concern are carbon dioxide, heavy hydrocarbons, and sulfur-bearing compounds. The greater the molecular weight of a compound, the greater its adsorption potential.

7.2.6.2 Temperature

7.2.6.2.1 General Considerations

Operation is very sensitive to the temperature of the incoming gas. Efficiency decreases as the temperature increases. Molecular sieves and most other adsorbents have significantly higher adsorptive capacity at low temperatures.

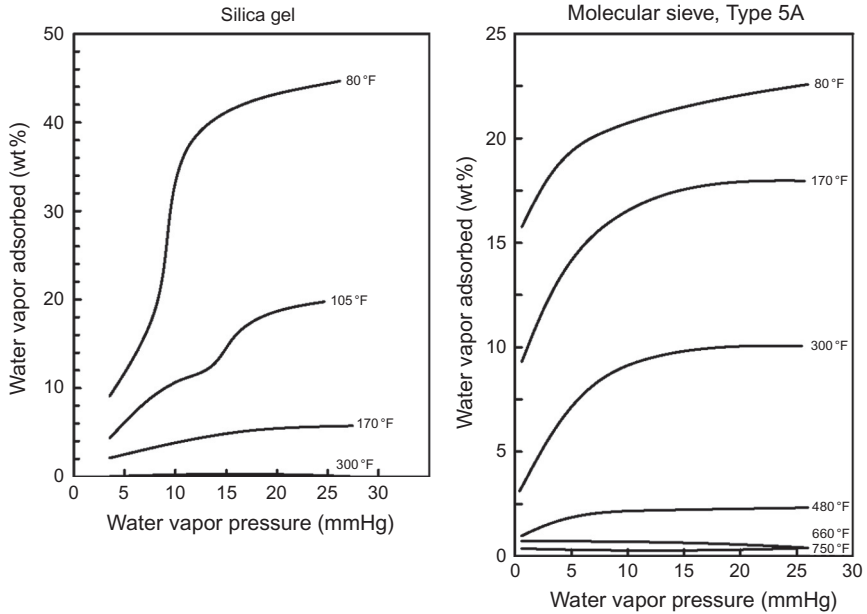


Figure 7.6 Effect of temperature on adsorption capacity.

Figure 7.6 indicates this characteristic for both silica gel and a type 5A molecular sieve. Water capacity of silica gel at 80 °F increases to more than twice that for the molecular sieve at higher water partial pressures. The temperature of the regeneration gas that commingles with the incoming wet gas ahead of the dehydrators is important. The temperature must remain within 10–15 °F, otherwise liquid water and hydrocarbons will condense as the hotter gas stream cools. Condensed liquids that strike the bed can shorten the solid desiccant's life.

The temperature of the hot gas entering and leaving a desiccant tower during the heating cycle affects plant efficiency and desiccant life. A high regeneration gas temperature ensures good removal (desorption) of water and contaminants from the bed.

7.2.6.2.2 Maximum Temperature

The maximum temperature depends on the type of contaminant and “holding power,” or affinity, of the desiccant for the contaminants. Typically, 450–600 °F is used. The desiccant bed temperature reached during the cooling cycle is important.

If wet gas is used to cool the desiccant, the cooling cycle should be terminated when the bed reaches 125 °F. Additional cooling may cause water to be adsorbed from the wet gas stream and preload (pre-saturate) the bed before the next adsorption cycle begins.

If dry gas is used to cool the desiccant, the cooling cycle should be terminated within 10–20 °F of the incoming gas temperature. This temperature maximizes adsorption capacity of the bed.

The temperature of the regeneration gas going through the regeneration gas scrubber should be held low enough to condense and remove the water and hydrocarbons without causing hydrate problems.

7.2.6.3 Pressure

The adsorption capacity of a dry bed unit decreases as pressure is lowered and with usage. Operating dry bed dehydrators well below the design pressure requires the desiccant to work harder to remove the additional water and maintain the desired effluent dew point.

With the same volume of incoming gas, the increased gas velocity occurring at the lower pressure could affect the effluent moisture content and damage the desiccant. At pressures above 1300–1400 psia, the co-adsorption effects of hydrocarbons are very significant.

7.2.6.4 Cycle Time

Most adsorbers operate on a fixed drying cycle time that is frequently set for the worst conditions. Adsorbent capacity is not a fixed value and declines with usage. For the first few months of operation, a new desiccant normally has a high capacity for water removal. If a moisture analyzer is used on the effluent gas, a much longer drying cycle can be achieved.

As the desiccant ages, the cycle time can be shortened to save regeneration fuel costs and improve the desiccant life. Common cycle times are:

- 8 h on stream
- 5–6 h heating
- 2–3 h cooling

7.2.6.5 Gas Velocities

As the gas velocity during the drying cycle decreases, the ability of the desiccant to dehydrate the gas increases by lowering effluent moisture contents and by longer drying cycle times. [Figure 7.7](#) shows the general effect of gas rate on the extent of dehydration.

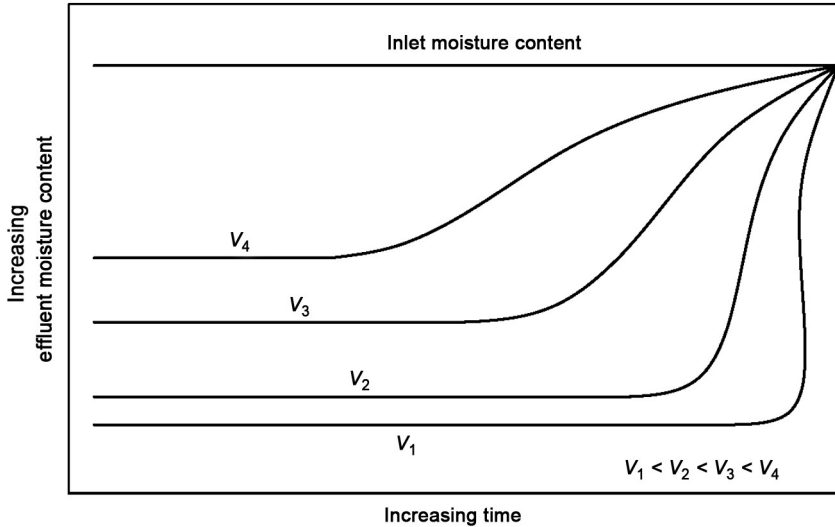


Figure 7.7 Series of adsorption curves that show effects of flow rate on the adsorption capacity of a single desiccant.

On the surface, it would seem desirable to operate at minimum flow rates to utilize the desiccant fully. However, low linear velocities require towers with large cross-sectional areas to handle a given gas flow and allow wet gas to channel through the desiccant bed and thus not be properly dehydrated.

Compromise must be made between the tower diameter and the maximum utilization of the desiccant, as illustrated in Figure 7.8. Maximum superficial velocities are shown in Table 7.1.

High linear velocities lower adsorption efficiency and may cause desiccant damage. The minimum tower diameter can be determined from the following:

$$d^2 = 3600 \left(\frac{Q_g T Z}{V P} \right) \quad (7.1)$$

where d = tower internal diameter, in.; Q_g = gas flow rate, MMSCFD; T = gas temperature, °R; Z = compressibility factor; V = gas superficial velocity, ft./min (Table 7.1); P = tower operating pressure, psia.

The regeneration gas velocity is important, especially when effluent moisture contents below 1 ppm are needed. At velocities <10 ft./s, hot gas will channel through the bed, leaving excess water in the bed after regeneration, which results in poor dehydration.

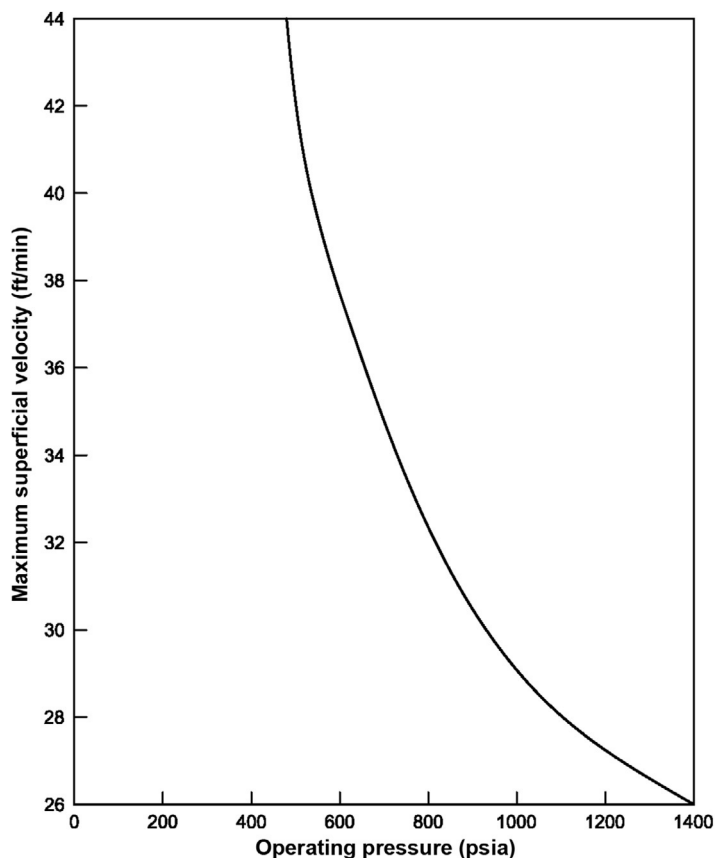


Figure 7.8 Maximum downflow of gas velocity as a function of operating pressure.

Table 7.1 Maximum Superficial Velocities

Tower Operating Pressure (psig)	Max. Superficial Velocity (ft./min)
14.7	110
400	60
600	55
1000	40

7.2.6.6 Source of Regeneration Gas

Source of regeneration gas depends on plant requirements and availability of a suitable gas stream. Regeneration gas should be dry when low effluent moisture contents (in the range of 0.1 ppm) are required. Plant tail gate gas can normally be used. If only moderate drying is required, a portion of the wet feed gas can be used.

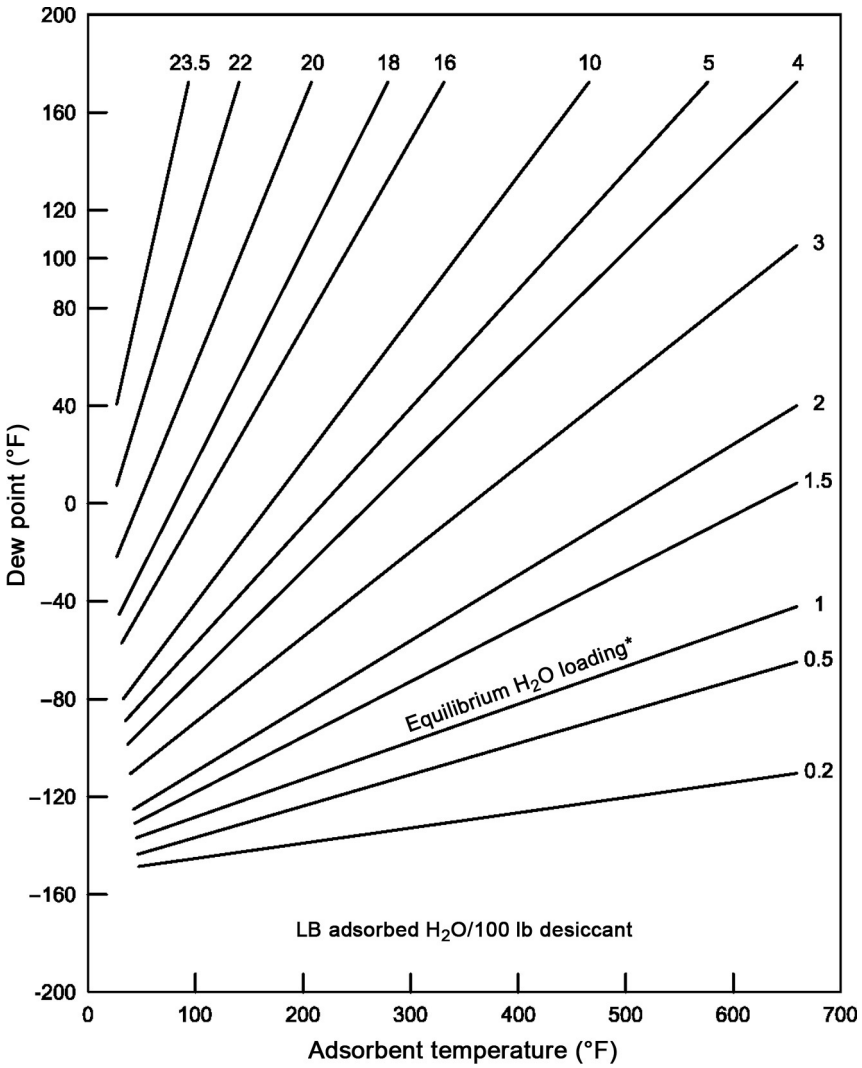


Figure 7.9 Equilibrium diagram showing lines of constant water loading for a type 4a molecular sieve.

Figure 7.9 is an equilibrium diagram showing lines of constant water loading. For example: a molecular sieve bed at 100 °F in equilibrium with a gas having a -80 °F water dew point will contain about 4 wt% water. Equilibrium curves for a given adsorbate-adsorbent can be used to estimate the regeneration conditions necessary to provide the required outlet conditions. For example, if the regeneration gas is taken from inlet gas with a dew

point of 40 °F and is heated to 450 °F, the mol sieve will contain 3 wt% water after regeneration. If the gas to be treated is at 100 °F, the intersection of the 3 wt% line with an adsorbent temperature of 100 °F gives the minimum attainable dew point at -95 °F. If this dew point is not satisfactory, either the regeneration gas must be heated to above 450 °F or a gas of a higher dew point (e.g., residue gas) must be used for regeneration gas.

7.2.6.7 Direction of Gas Flow

Flow direction influences effluent purity, regeneration gas requirements, and desiccant life. The direction of flow during the drying cycle is downward, which permits higher velocities without lifting or fluidizing the desiccant bed and means fluidization can severely damage the desiccant.

Direction of flow during the heating cycle is counter-current to the direction of the adsorption flow. It permits better reactivation of the lower portion of the desiccant bed, which must perform the super-dehydration during the drying cycle, especially in cryogenic plants. If flow is co-current, all water and/or other contaminants must move through the entire bed, thus causing additional desiccant contamination and requiring longer regeneration times.

Direction of flow during the cooling cycle depends on whether dry or wet gas is used. When dry gas is used, the flow direction is counter-current to the adsorption flow, thus simplifying piping and valve configuration. When wet gas is used, the flow direction is in the same direction as the adsorption flow so that the water adsorbed during the cooling cycle as the desiccant cools will preload on the inlet end of the bed.

If counter-current flow is used, water is deposited on the exit end of the bed. When the next adsorption cycle begins, the wet gas is immediately dried. As the dry gas continues to move down through the bed, it picks up some of the water deposited during the cooling cycle and sometimes puts too much moisture in the effluent stream. If wet gas is used, the additional water load, deposited during the cooling cycle, should be included when the amount of desiccant needed for dehydration is calculated.

7.2.6.8 Desiccant Selection

No desiccant is best for all applications. Desiccant selection is based on economics and process conditions. Desiccants are usually interchangeable. Equipment designed for one desiccant can often operate effectively with another. No desiccant product will remain effective with massive liquid carryovers.

All desiccants are damaged by heavy impurities carried into the bed with gases. These include crude oil and condensate, glycols and amines, most corrosion inhibitors, and well-treating fluids. All desiccants exhibit a decrease in capacity (design loading) with an increase in temperature. Molecular sieves are less affected while aluminas are most affected. Aluminas and molecular sieves act as a catalyst with H_2S to form COS , which deposits sulfur on the desiccant bed during regeneration. Alumina gels, activated aluminas, and molecular sieves are all chemically attacked by strong mineral acids and thus their adsorptive capacity is decreased. Special acid-resistant molecular sieves desiccants are available.

Table 7.2 provides certain physical characteristics of the more common solid desiccants.

7.2.6.8.1 Molecular Sieves

Molecular sieves offer the highest adsorptive capacity of all desiccants when the feed gas is at very high temperatures or at low relative saturation. They are the only desiccants capable of dehydrating gas to <1 ppm of water content, which is required for cryogenic temperatures (dew points down to -150°F).

7.2.6.8.2 Silica Gel and Alumina

Water-saturated gases entering the dehydrator can adsorb twice as much water as molecular sieves and offer a lower first cost.

7.2.6.8.2.1 Silica Gel Silica gel can be regenerated to a lower water content than molecular sieves and at much lower temperatures (400°F for

Table 7.2 Properties of Solid Desiccants

Desiccant	Bulk Density (lb/ft. ³)	Specific Heat (Btu/lb/°F)	Normal Sizes Used	Design Adsorptive Capacity (wt%)
Activated alumina	51	0.24	¼ in.– 8 mesh	7
Mobil SOR beads	49	0.25	4–8 mesh	6
Fluorite	50	0.24	4–8 mesh	4–5
Alumina gel (H-151)	52	0.24	⅜–¾ inch	7
Silica gel	45	0.22	4–8 mesh	7
Molecular sieves (4A)	45	0.25	⅜ inch	14

gels vs. 500–600 °F for sieves). It shatters in the presence of free water or light hydrocarbon liquids. The problem is minimized by using a 4–6 in. buffer bed of mullite ball (or equivalent) to protect the silica gel from direct contact.

7.2.6.8.3 Desirable Characteristics of Solid Desiccants

The desirable characteristics of solid desiccants are:

- High adsorptive capacity (lb/lb), which reduces contactor size
- Easy regeneration for simplicity and economics of operation
- High rate of adsorption, which allows higher gas velocities and thereby reduces contactor size
- Low resistance to gas flow, which minimizes gas pressure drop through the unit
- High adsorptive capacity retained after repeated regeneration, which allows smaller initial charge and longer service before replacement
- High mechanical strength to resist crushing and dust formation
- Inert chemicals to prevent chemical reactions during adsorption and regeneration
- Volume unchanged when product is wet, which would otherwise necessitate costly allowance for expansion
- Noncorrosive and nontoxic properties, which eliminates the necessity for special alloys and costly measures to protect the operator's safety
- Low cost, which reduces initial and replacement costs

7.2.6.9 Effect of Regeneration Gas on Outlet Gas Quality

Regeneration gas desorbs molecular sieve beds chromatographically in the reverse order of the adsorption bed. For example, adsorbed methane and ethane would be desorbed first, then propanes and heavier hydrocarbons, then carbon dioxide. This is followed by any hydrogen sulfide that might have been in the inlet gas and, lastly, by the water. The effect of the concentration of these impurities in the regeneration gas stream may be significant when regeneration gas is 10–15% of the net inlet gas.

In the regeneration circuit, the bulk of the water and some heavy hydrocarbons are condensed and removed from the system. These components may render the sales gas off-specification for a short period. The peak of ethane or CO₂ could cause the sales gas to exceed its heating value. Concentrations of 3–4 ppm of H₂S can be concentrated up to 20 times that amount, and thus render the composite stream far off-spec.

Figure 7.3 shows the cooled regeneration gas stream recombined with the main gas inlet to be processed. This recycle essentially eliminates the problem of making the sales gas off-specification. However, it adds cost because the main gas-processing capacity must be increased appropriately. If the sales gas limits are no problem, or if there is other downstream processing, the cooled, scrubbed regeneration gas may be admitted directly to the dried outlet gas without this recycle.

7.2.6.10 Pressure Drop Considerations

To achieve acceptable dehydration and extend the life of the desiccant, the pressure drop through the dehydration tower should not exceed 8 psi. Pressure drop through the tower can be estimated from either desiccant pressure drop curves furnished by the manufacturer (Figure 7.10), or pressure drop equation.

Pressure drop through a desiccant tower can be estimated from the following equation:

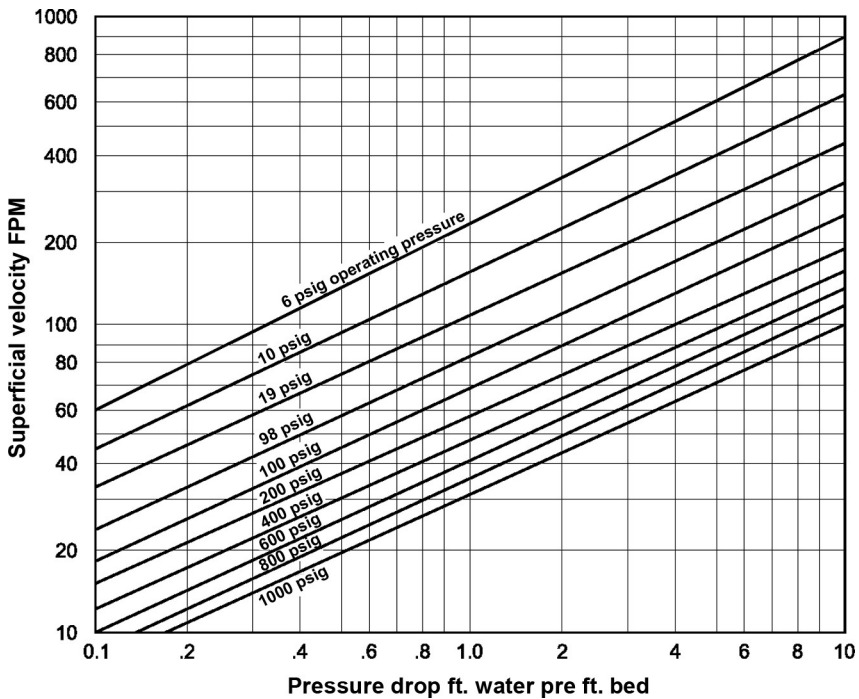


Figure 7.10 Typical pressure-drop curve for silica gel type desiccants, 0.15-in. diameter beads.

Table 7.3 Constants Used in Pressure Drop Equations

Particle Type	B	C
1/8-inch bead	0.0560	0.0000889
1/8-inch extruded	0.0722	0.0001240
1/16-inch bead	0.1520	0.0001360
1/16-inch extruded	0.2380	0.0002100

$$\frac{\Delta P}{L} = B\mu V + C\rho V^2 \quad (7.2)$$

where ΔP = pressure drop through the tower, psi (typically sized for 5 psi); μ = gas viscosity, cp; ρ = gas density, lb/ft.³; V = gas superficial velocity, ft./min; B , C = constants provided in Table 7.3.

Example 7.1 Determination of Pressure Drop Through a Dry Bed Desiccant Dehydration Tower

Given

- Superficial gas velocity = 40 ft./min
- Tower operating pressure = 1000 psig
- Gas molecular weight = 18
- Bed height (L) = 30 ft.
- Desiccant type = silica gel
- Desiccant diameter = 0.15 in.
- Desiccant pressure drop curve (Figure 7.10)

Notes

- Curves are based on air flow. For other gases, multiply pressure drop by

$$\left(\frac{MW_{\text{Gas}}}{MW_{\text{Air}}} \right)^{0.9} \quad (7.3)$$

- Pressure drop curves are based on clean beds. After about 2 years, the beds will foul somewhat and the pressure drop will be about 1.6 times the value read from the curves.

Solution

- Enter Figure 7.10, extend a horizontal line from superficial velocity of 40 ft./min, and intersect with the operating pressure of 1000 psig.
- Draw a vertical straight line down from the intersection and read a pressure drop of 1.9 ft. of water per foot of bed.
- Calculate the total pressure drop across the bed after 2 years of service,

$$\begin{aligned} \text{Total } \Delta P &= \left(1.9 \frac{\text{Ft. of H}_2\text{O}}{\text{Ft. of Bed}} \right) \left(0.433 \frac{\text{psi}}{\text{Ft. of H}_2\text{O}} \right) \left(\frac{18}{29} \right)^{0.9} (1.6)(30 \text{ ft.}) \\ &= 25 \text{ psi} \end{aligned}$$

7.2.7 Equipment

The proper selection of equipment is essential to good operations.

7.2.7.1 Inlet Gas Cleaning Equipment

All hydrocarbon liquids, free water, glycol, amine, or lube oil carryover must be cleaned from the inlet gas to ensure the best dry desiccant dehydrator operation. In all cases, the dry bed unit should have a scrubber (or a filter separator) between it and a primary well fluid separator.

A microfiber filter separator (or its equivalent) should always be installed upstream of the inlet scrubber if a carryover of glycols, amines, or compressor lube oils is possible.

Liquid level controls need to be checked frequently as well as the liquid dump line to ensure their operability.

7.2.7.2 Adsorber Tower

7.2.7.2.1 General Considerations

An adsorber tower is a cylindrical tower filled with a solid desiccant. The depth of desiccant will vary from a few feet to 30 ft. or more. The vessel diameter may be as much as 10–15 ft. or more. The bed height to diameter (L/D) ratio of 2.5–4.0 to 1 is desirable. Lower ratios (1:1) are sometimes used, which could result in poor gas dehydration caused by nonuniform flow, channeling, and inadequate contact time between the wet gas and the desiccant.

Three problems that frequently cause poor operation are insufficient gas distribution, inadequate insulation, and improper bed supports.

Figure 7.11 illustrates many of the desired features recommended in a dry bed desiccant tower.

7.2.7.2.2 Insufficient Gas Distribution

Poor gas distribution at the inlet and outlet of the desiccant beds has caused many costly problems, resulting in channeling and desiccant damage. The inlet gas distributor should be provided with adequate baffling before the gas enters the desiccant bed. A void space of 18–24 in. is recommended. Neither gas to be dehydrated nor the regeneration gas should impinge directly on the bed.

Channeling, high localized velocities, and swirling can cause desiccant attrition and high-pressure drop through the desiccant bed as attrition fines lodge between the regular particles. Screen-wrapped slotted pipe, with gas at low velocities exiting radially into the vessel is recommended. A 4- to 6-in. layer of large diameter (2 in.) support balls can be placed on top of the

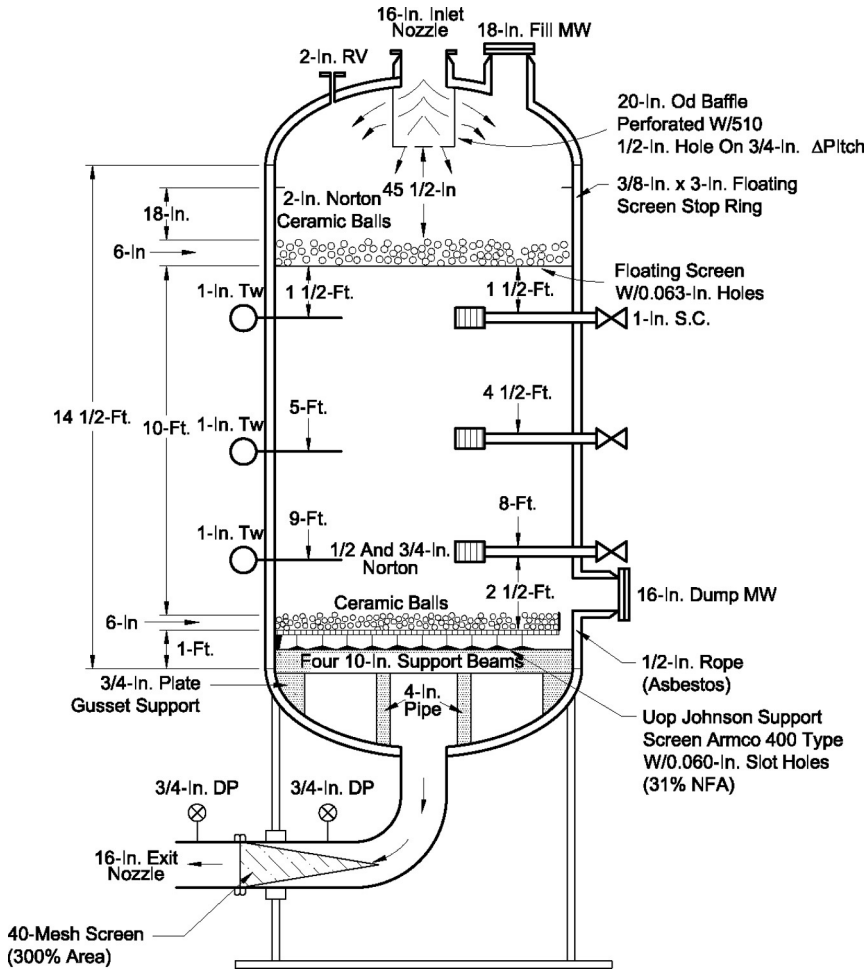


Figure 7.11 Molecular sieve gas dehydration tower.

desiccant bed. This improves gas distribution and prevents desiccant damage from swirling. Swirling can destroy several feet of castable refractory lining by turning the powdered desiccant into a sandblasting agent, which results in high heat losses and poor desiccant regeneration.

7.2.7.2.3 Inadequate Insulation

Internal or external insulation can be used. Internal insulation reduces the total regeneration gas requirements and costs and eliminates the need to heat and cool adsorber vessels. Provisions must be made for expansion and

contraction so that there will be no cracking or weld failures. Vessels are normally made from a castable refractory lining. Liner cracks permit some of the wet gas to bypass the desiccant bed. Only a small amount of wet, bypass gas can cause freeze-up in cryogenic plants. Ledges installed every few feet along the vessel wall can help eliminate liner cracks.

7.2.7.2.4 Improper Bed Supports

Two common bed supports include horizontal screens supported by I-beams and a welding ring and vessel whose bottom head is filled with graduated support balls. Screens are usually made of stainless steel or monel that have openings at least 10 meshes smaller than the smallest desiccant particle. A 0.033-in. slot opening will retain standard desiccant particles.

Screen traps crush desiccant particles, which prevents the malfunction of downstream equipment caused by invasion of these particles as the desiccant deteriorates. Screens should be securely fastened in the vessel.

Provisions should be made for expansion and contraction as the adsorbers heat and cool. Annular space between the vessel wall and the edge of the bed support screen must be sealed to prevent the loss of desiccant. Asbestos rope packing, forced in this space, has been used. A support ring around the edges of the screen is beneficial. If the screens are installed in sections, they should be fastened securely with stainless steel wire.

Support balls on the screens are helpful. Two to three inches of ½-in. balls are gently placed on the screen and a 2 or 3 in. smooth layer of ¼-in. balls is gently placed on top of the ½-in. balls. These layers prevent desiccant dust or whole particles from plugging the screen openings and forcing a high-pressure drop across the desiccant beds. When calculating the regeneration needs of the system, it is important to include the heat requirements for the support balls.

If the bottom head of the vessel is filled with graduated support balls, a gas distributor may be required between the balls and the lower portion of the desiccant bed when upflow heating or cooling is used. This is important on large-diameter vessels to prevent channeling and poor reactivation of the desiccant. Many adsorbers have a void area in the bottom, below the bed supports, to collect contaminants, dust, and fines. A blowdown nozzle can be provided to discharge these materials.

A moisture sample probe should be located in the adsorbers in cryogenic plants several feet from the outlet end of the bed and extending to the center. This probe, used in conjunction with the outlet gas moisture probe, offers valuable flexibility in studying and solving dehydrator problems, particularly

for determining if gas is being channeled down the walls of the vessel. It permits capacity tests for optimizing drying cycle times.

Tests can be conducted with reasonable safety because movement of the water front can be detected prior to breakthrough. The probe can be a long thermowell drilled with 1/32-in. holes on the sides near the end of the probe.

7.2.7.2.5 Pressurization

For the best performance and maintenance of desiccant quality, adsorbers should never be pressurized faster than 50 psi/min and should never be depressurized faster than 10 psi/min. Downflow pressure drop should not exceed 1 psi/ft. Upflow pressure drop should not be $<1/4$ psi/ft. to prevent fluff fluidization.

Even with the best designs, some desiccant dust is swept out of the beds at design gas-flow rates. Certain amounts can be tolerated in many field dehydration systems. It is not acceptable in turbo-expander plant designs that involve extensive downstream heat exchange and processing. A regenerative gas heater is sized to provide the heat to desorb the water and the heat to heat the absorber shell. Temperatures range between 500 to 550 F. In many instances, this problem can be solved with microfiber filters (cleaning to 1 μm) with a differential pressure of 15 psi.

7.2.7.2.6 Regeneration Gas Exchangers, Heaters, and Coolers

A gas or gas exchanger is usually designed with the following assumptions: All of the water will be liberated from the bed in 1 h at 250 °F, and regeneration gas can be cooled to within 10 °F of the sales gas temperature. The heat of desorption for silica gel is 1100 Btu/lb of water and for a molecular sieve it is ~50% higher.

Heat required to heat the desiccant can be calculated using the following equation:

$$Q = WC_p \Delta t \quad (7.4)$$

where Q =heat required, Btu; W =weight of desiccant, lb; C_p =specific heat of desiccant, Btu/lb/°F; Δt =difference in desired bed temperature and normal bed operating temperature, °F.

Sensible heat for the contactor shell can be calculated using Equation (7.4) and by estimating the weight of the steel using 0.12 Btu/lb/°F for C_p . On units that have internal insulation, the heat transferred to the shell

is considered negligible. Normal practice is to add 10–20% to the sum of the heat required to account for heat losses and to add some margin of safety.

7.2.7.2.7 Regeneration Gas Separator

Most desiccants also have an affinity for hydrocarbons, thus a skimmer is used to separate the valuable hydrocarbons from the water to be discarded. Frequent pH tests on the discarded water helps pinpoint corrosion problems in the adsorption system.

A common problem encountered in regeneration gas separators is the fouling of the liquid dump line by desiccant dust and heavy oils. This allows liquids to be carried back where they can damage the desiccant in a recycle system or contaminate sales and downstream facilities in a once-through system. Regular inspection and cleaning are required to prevent such damage.

7.2.7.2.8 Control Valves

Quality valves should be used to prevent costly operating problems. Generally, two-way valves have fewer problems than three-way valves. Most difficult service is encountered when the valves have hot regeneration gas (600 °F) on one side and ambient (100 °F) gas on the other side. Careful piping design can reduce this large gradient. Valve sequencing is important to prevent a sudden upflow caused by a pressure difference. This problem could fluidize the bed and damage the desiccant.

Dry bed dehydrators equipped with motor valves for switching operations require frequent servicing to eliminate leakage.

7.2.8 Expander Plant Molecular Sieve Applications

Turbo-expander plants commonly operate down to temperatures of -150 °F.

Operating points:

Turbo-expander plants operate much below the equilibrium water content data illustrated in McKetta-Wehe chart.

They include designs to water contents as low as 1 ppm. As shown in [Table 7.2](#), only molecular sieves and activated alumina are capable of such performance.

Molecular sieves are used in ~95% of the dehydration equipment for this type of plant (a 4A molecular sieve has twice the adsorptive capacity of activated alumina).

[Figure 7.12](#) compares the adsorptive capacity of several desiccants at lower relative humidities of the gas.

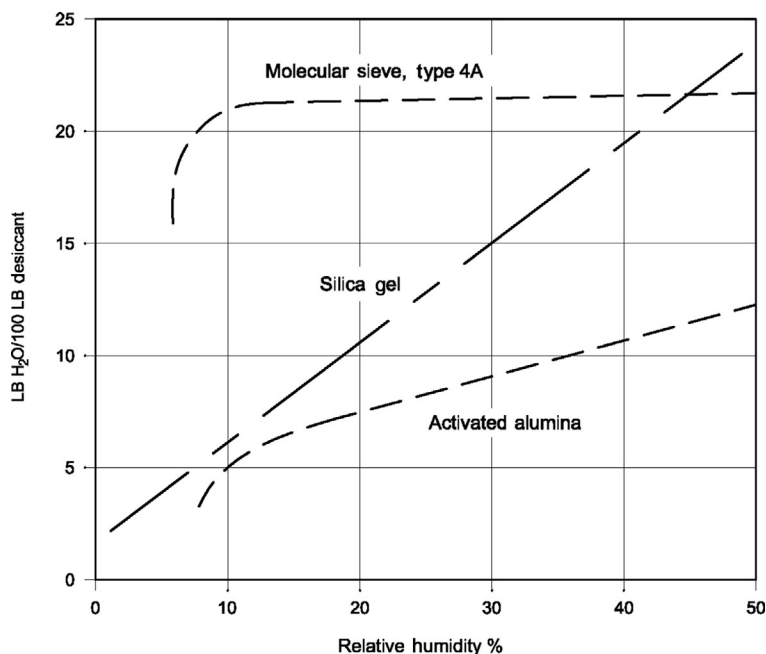


Figure 7.12 Water vapor adsorption at 60 °F.

At 30% relative humidity, molecular sieves would adsorb 21.5 lb water per 100 lb of desiccant, whereas silica gel would absorb 15 lb of water/100 lb of desiccant.

7.2.9 Desiccant Performance

7.2.9.1 General Conditions

Desiccants decline in adsorptive capacity at different rates under varying operating conditions. Desiccant aging is a function of many factors, including number of cycles experienced and exposure to any harmful contaminants present in the inlet stream that are not completely removed during normal reactivation.

The single most important variable affecting the decline rate of desiccant capacity is the chemical composition of the gas or liquid to be dried. The feed stream composition should always include the contaminants. The capacity of a new desiccant will decline slowly during the first few months in service because of cyclic heating, cooling, and netting. Desiccant capacity usually stabilizes at about 55–70% of the initial capacity.

7.2.9.2 Moisture Analyzer

A moisture analyzer is used to optimize the drying cycle time. It allows drying time to be shortened as the desiccant ages. Both inlet and outlet moisture analyzer probes should be used. A probe extending ~2 ft. upward into the bed from the outlet end is recommended because it allows a dehydration capacity test to be run without the risk of a water breakthrough.

7.2.9.3 Effect of Contaminants in Inlet Feed Stream

Compressor oils, corrosion inhibitors, glycols, amines, and other high-boiling contaminants cause a decline in desiccant capacity, because normal reactivation temperatures will not vaporize the heavy materials. Residual contaminants slowly build up on the desiccant's surface reducing the area available for adsorption. Many corrosion inhibitors chemically attack certain desiccants, permanently destroying their usefulness.

7.2.9.4 Effect of Regeneration Gases Rich in Heavy Hydrocarbons

Use of this rich gas in a 550–600 °F regeneration service aggravates coking problems. Rich gases may be dried satisfactorily with molecular sieves. Lean dry gas is always preferable for regeneration.

7.2.9.5 Effect of Methanol in the Inlet Gas Stream

Methanol in the inlet gas is a major contributor to the coking of molecular sieves where regeneration is carried out at temperatures above 550 °F. Polymerization of methanol during regeneration produces dimethyl ether and other intermediates that will cause coking of the beds. Conversion to ethylene glycol injection, instead of methanol for hydrate control, will increase sieve life and add at least 10% to sieve capacity by removing the vapor phase methanol from the system.

7.2.9.6 Useful Life

The useful life of a desiccant ranges from 1 to 4 years in normal service. Longer life is possible if feed gas is kept clean. The effectiveness of regeneration plays a major role in retarding the decline of a desiccant's adsorptive capacity and prolonging its useful life. If all the water is not removed from the desiccant during each regeneration, its usefulness will sharply decrease.

7.2.9.7 Effect of Insufficient Reactivation

Insufficient reactivation can occur if the regeneration gas temperature or velocity is too low. A desiccant manufacturer will generally recommend

the optimum regeneration temperature and velocity for the product. The velocity should be high enough to remove the water and other contaminants quickly, thus minimizing the amount of residual water and protecting the desiccant.

7.2.9.8 Effect of High Reactivation Temperature

Higher reactivation temperatures remove volatile contaminants before they form coke on the desiccant. Higher temperature maximizes desiccant capacity and ensures minimum effluent moisture content. The final effluent hot gas temperature should be held one or two hours to achieve effective desiccant reactivation.

7.2.10 Areas Requiring Engineering Attention

The design of adsorption type dehydration equipment can be improved by considering the major process variables that affect bed loading and proper conditioning of inlet gas and proper design of regeneration gas systems. An accurate estimation of bed sizes is necessary to realistically evaluate the competitive offerings of desiccant vendors. Improvement in the design of adsorber internals, including internal insulation, improved switching valves, and control systems should also be evaluated.

Example 7.2 Preliminary Dry (Solid) Bed Desiccant Design

Note

Detailed design of dry bed dehydrators should be left to experts. The general “rule of thumb” presented herein can be used for preliminary design.

Given

Feed rate = 50 MMSCFD

Molecular weight of gas = 17.4

Operating temperature = 110 °F

Operating pressure = 600 psia

Inlet dew point = 100 °F (equivalent to 90 lb H₂O/MMSCF)

Desired dew point = 1 ppm H₂O

Gas density = 1.70 lb/ft.³

Gas Analysis

Component	Mole Percent
N ₂	4.0
C ₁	92.0
C ₂	2.4
C ₃	0.3
iC ₄ +	<u>1.0</u>
	100.0

Determine

Design a dry desiccant dehydrator.

Solution

1. Water adsorbed

For this example, an 8-h on-stream cycle with 6 h of regeneration and cooling will be assumed. On this basis, the amount of water to be adsorbed per cycle is

$$= \left(\frac{8}{24} \right) (50 \text{ MMSCF}) \left(90 \frac{\text{lb}}{\text{MMSCF}} \right)$$

$$= 1500 \text{ lb H}_2\text{O/cycle}$$

2. Loading

Because of the relative high operating temperature, use Mobil's SOR beads as the desiccant and design on the basis of 6% loading. SOR beads weigh $\sim 49 \text{ lbs/ft.}^3$ (bulk density) (refer to [Table 7.2](#)).

The required weight of desiccant per bed is:

$$= \frac{1500 \text{ lb H}_2\text{O}}{(0.06 \text{ lb H}_2\text{O/lb Desiccant})}$$

$$= 25,000 \text{ lb desiccant per bed}$$

The required volume of desiccant per bed is:

$$= \frac{25,000 \text{ lb desiccant per bed}}{49 \text{ lb desiccant/ft.}^3}$$

$$= 510 \text{ ft.}^3/\text{per bed}$$

3. Tower sizing

Recommended maximum superficial velocity at 600 psia is about 55 ft./min. ([Table 7.1](#)).

Minimum vessel internal diameter (from Equation 7.1)

$$d^2 = 3600 \left(\frac{Q_g TZ}{VP} \right)$$

$$d^2 = 3600 \frac{(50)(570)(1.0)}{(55)(600)}$$

$$d = 55.7 \text{ in. or } 4.65 \text{ ft.}$$

Bed height is

$$L = \frac{50 \text{ ft.}^2}{\frac{\pi (4.65)^2}{4 \text{ ft.}^2}} = 30 \text{ ft.}$$

The pressure drop across a clean bed, assuming $\frac{1}{8}$ -in bead and $\mu = 0.01$ cp, (Equation 7.2) is

$$\begin{aligned}\Delta P &= |B\mu V = C_p V^2|L \\ \Delta P &= ((0.056)(0.01)(55) + (0.00009)(1.70)(55)^2)30 \\ &= 14.8 \text{ psi}\end{aligned}$$

This is higher than the maximum recommended pressure drop of 8 psi, thus the vessel internal diameter should be increased to the next standard size.

Choose a diameter of 5 ft. 6 in., substitute into the above equations and determine V , L , and ΔP .

$V = 39.2$ ft./min; $L = 21.5$ ft.; $\Delta P = 5.5$ psi.

Allowing 6 ft. space to remove the desiccant and refill would be about 28 ft. This yields an L/D of $28/5.5 = 5.0$, which is acceptable.

4. Regeneration heat requirement

Assume the bed (and tower) is heated to 350°F . The average temperature will be $(350 + 110)^\circ\text{F}/2 = 230^\circ\text{F}$.

The approximate weight of 5 ft. 6 in. ID \times 28 ft. \times 700 psig tower is 53,000 lbs including the shell, heads, nozzles, and supports for the desiccant.

Heating and cooling requirement can be determined using Equation (7.4)

$$Q = wC_p\Delta T$$

where Q = heat required, Btu; C_p = specific heat of material at average temperature, Btu/lb/ $^\circ\text{F}$; Δt = temperature differential across material, $^\circ\text{F}$.

5. Heating requirement/cycle

Desiccant: (25,000 lb)

$(350 - 100^\circ\text{F})(0.25)^{(4)} = 1,500,000$ Btu

Tower: (53,000 lb)

$(350 - 100^\circ\text{F})(0.12)^{(1)} = 1,520,000$ Btu

Desorb water: (1500 lb)

$(1100 \text{ Btu/lb})^{(2)} = 1,650,000$ Btu

$(1500 \text{ lb})(230 - 110^\circ\text{F})(1.0)^{(3)} = 200,000$ Btu

Total heat = 4,870,000 Btu

10% heat losses = 490,000 Btu

Total heat requirement/cycle = 5,360,000 Btu/cycle

Notes:

(1) Specific heat of steel.

(2) The number "1100 Btu/lb" is the heat of water desorption, a value supplied by the desiccant manufacturer.

- (3) The majority of the water will desorb at the average temperature. This heat requirement represents the sensible heat required to raise the temperature of the water to the desorption temperature.

- (4) Specific heat of SOR beads “R” (refer to [Table 7.2](#)).

6. Cooling requirement/cycle

Desiccant: (25,000 lb)

$$(350 - 100^\circ\text{F})(0.25)^{(3)} = 1,500,000 \text{ Btu}$$

Tower: (53,000 lb)

$$(350 - 100^\circ\text{F})(0.12)^{(1)} = 1,520,000 \text{ Btu}$$

Total cooling = 3,020,000 Btu

10% for nonuniform cooling = 300,000 Btu

Total cooling requirement/cycle = 3,320,000 Btu/cycle

This example assumes insulation is on the outside of the towers. The duty would be less if the towers were insulated internally. Internal insulation should be used to minimize thermal stress caused by wide swings in temperature during regeneration. Channeling, or bypassing, gas around the desiccant beds can be a problem.

7. Regeneration gas heater

Assume inlet temperature of regeneration gas is 400°F . The initial outlet temperature of the bed will be the bed temperature of 110°F ; at the end of the heating cycle, the outlet temperature will be the design value of 350°F . Therefore, the average outlet temperature is $(350 + 110) / 2 = 230^\circ\text{F}$.

The volume of gas required for heating will be

$$V_{\text{heating}} = \frac{5,360,000 \frac{\text{Btu}}{\text{cycle}}}{(400 - 230)^\circ\text{F}(0.64) * \text{Btu/lb}/^\circ\text{F}}$$

$$= 49,400 \text{ lbs} = \text{cycle}$$

The regeneration gas heater load, Q_H , is then:

$$Q_H = 49,400(400 - 11)(0.62) * \text{Btu/lb}/^\circ\text{F}$$

$$= 8,900,000 \text{ Btu/cycle}$$

For design, add 25% for heat losses and nonuniform flow. Assuming a three-hour heating cycle, the regeneration gas heater must be sized for

$$Q_H = (8,900,000) \left(\frac{1.25}{3} \right)$$

$$= 3,710 \text{ Btu/h}$$

8. Regeneration gas cooler

Regeneration gas: 49,400

$$(230 - 110)(0.61)/3 = 1,205,000 \text{ Btu/h}$$

Water: $1500(1157 - 78)^{(1)}/0.5 = 3,237,000 \text{ Btu/h}$

Total load = 4,442,000 Btu/h

10% heat loss = 44,000 Btu/h

Total = 4,886,000 Btu/h

9. Cooling cycle

Similarly for the cooling cycle where the initial outlet temperature is 350 °F and at the end of the cooling cycle, it is ~110 °F. The average outlet temperature is $(350 + 110)/2 = 230 \text{ °F}$.

Assuming the cooling gas is at 110 °F, the volume of gas required for cooling will be

$$V_{\text{cooling}} = \frac{3,320,000 \text{ Btu/cycle}}{(230 - 100)^\circ\text{F}(0.59) * \text{Btu}/^\circ\text{F}}$$

$$= 46,900 \text{ lbs} = \text{cycle}$$

Notes: (1) from steam tables.



7.3 ABSORPTION

7.3.1 Process Overview

In the absorption process, a hygroscopic liquid is used to contact wet gas and remove the water vapor.

The most common liquid used in absorption type dehydration units is triethylene glycol (TEG).

7.3.2 Principles of Absorption

7.3.2.1 Absorption and Stripping

Through absorption, the water in a gas stream is dissolved in a relatively pure liquid solvent stream. The reverse process, in which the water in the solvent is transferred into the gas phase, is known as stripping. The terms regeneration, reconcentration, and reclaiming are also used to describe stripping (or purification) because the solvent is recovered for reuse in the absorption step. Absorption and stripping are frequently used in gas processing, gas sweetening, and glycol dehydration.

7.3.2.2 Raoult's and Dalton's Laws

Absorption can be qualitatively modeled by using Raoult's and Dalton's Laws. For a vapor liquid equilibrium system, Raoult's Law states that the partial pressure of a component in a vapor phase that is in equilibrium with

a liquid is directly proportional to the mole fraction of the component in the liquid phase. Dalton's Law states that the partial vapor pressure of a component is equal to the total pressure multiplied by its mole fraction in the gas mixture.

Raoult's Law expressed in equation form is:

$$p_i = P_i X_i \quad (7.5)$$

Dalton's Law expressed in equation form is:

$$p_i = P Y_i \quad (7.6)$$

where p_i = partial vapor pressure of component i ; P_i = vapor pressure of pure component i ; X_i = mole fraction of component i in the liquid; P = total pressure of the gas mixture; Y_i = mole fraction of component i in the vapor.

Combining these laws we have:

$$P Y_i = p_i X_i$$

or

$$p_i | P = Y_i | X_i \quad (7.7)$$

Since the pure-component vapor pressure and the total pressure are not affected by composition Equation (7.7) is significant. It states that the ratio of the vapor mole fraction to the liquid mole fraction for any component is independent of the concentrations of that component and the other components present. The ratio Y_i/X_i is commonly known as the K -value.

Since the pure-component vapor pressure increases with temperature, the K -value increases with increasing temperature and decreases with increasing pressure. In physical terms, this means the transfer from the gas phase to the liquid phase (absorption) is more favorable at lower temperature and high pressures. The transfer to the gas phase (stripping) is more favorable at higher temperatures and lower pressures.

7.3.3 Glycol-Water Equilibrium

Absorption processes are dynamic and continuous. Gas flow cannot be stopped to let the vapor and liquid reach equilibrium. Thus, the system must be designed to approach equilibrium as closely as possible while flow continues. This is accomplished by using a trayed or packed contactor in which the gas and liquid are in counter-current flow.

The closer to 100% equilibrium that a trayed or packed section approaches, the higher the tray or packing efficiency. For example, a common tray efficiency is 25%, meaning that 25% of the water molecules that would have been transferred under equilibrium conditions were actually transferred. Wet gas enters the bottom of the column and contacts the rich glycol (high water content) just before the glycol leaves the column. The gas encounters leaner glycol as it works its way up the column, contacting the leanest glycol (lowest water content) just before it leaves the column. Equilibrium is based on Dalton's and Raoult's laws and can be rearranged as follows:

$$Y_i = X_i \left(\frac{P_i}{P} \right) \quad (7.8)$$

Since P_i/P is constant for constant temperature, the concentration of the water in the gas must be directly proportional to the concentration in the liquid. However, the liquid concentration is constantly changing as water is absorbed. The counter-current flow in the contactor makes it possible for the gas to transfer a significant amount of water to the glycol and still approach equilibrium with the leanest glycol concentration.



7.4 GLYCOL DEHYDRATION

7.4.1 Principles of Operation

7.4.1.1 Introduction

After the liquid (free) water has been removed from the gas stream by separation, 25–120 lbs of water per MMSCF of gas will remain, depending on the temperature and pressure of the gas. The warmer the inlet gas and the lower the pressure, the more water vapor the gas stream will contain (see [Figure 7.13](#)). Normally, between 20 and 115 lbs of water per MMSCF of gas must be removed before the required dew point of the gas is met.

The schematics in [Figures 7.14](#) and [7.15](#) show the flow through a typical glycol dehydration system. The glycol dehydration process can be discussed in two parts:

- Gas system ([Figure 7.14](#))
- Glycol system ([Figure 7.15](#))

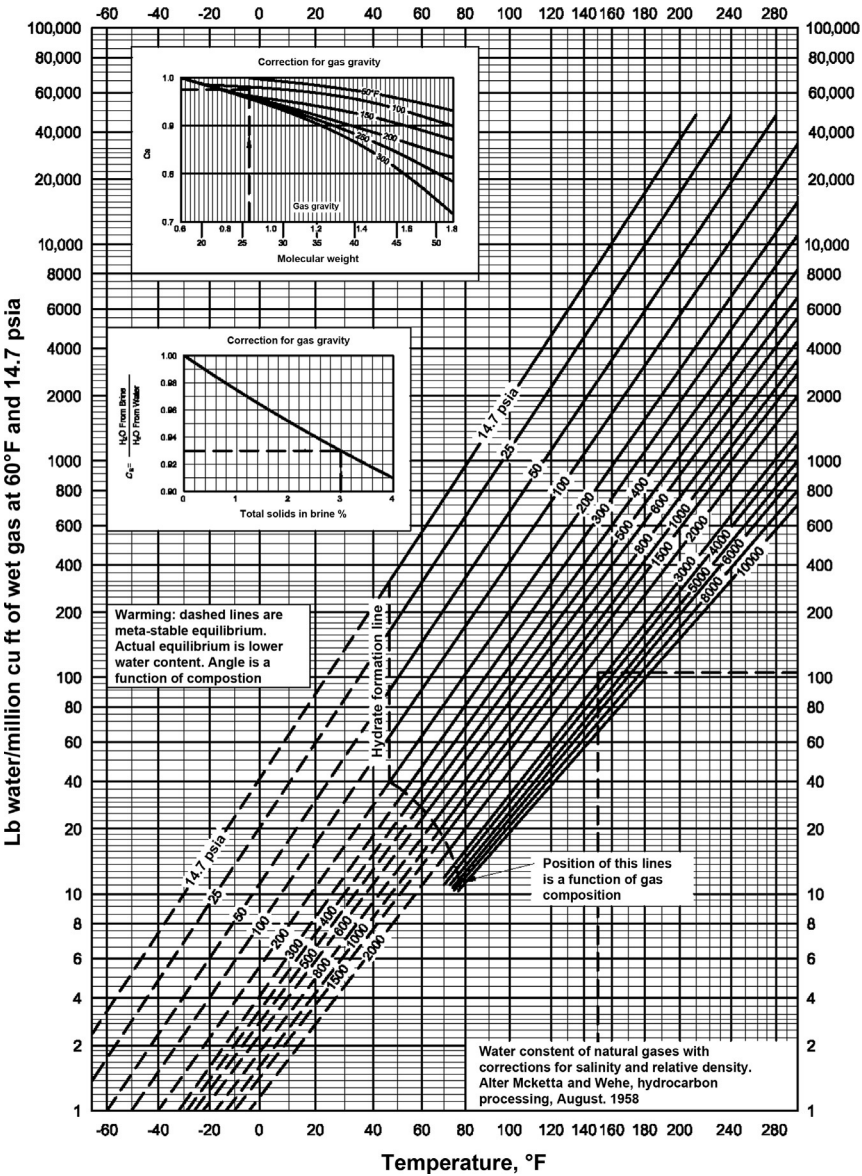


Figure 7.13 Water content of sweet, lean natural gas—McKetta-Wehe.

7.4.2 Gas System

7.4.2.1 Inlet Scrubber/Microfiber Filter Separator

Wet gas enters the unit through the inlet gas scrubber/microfiber filter separator, usually vertical, to remove liquid and solid impurities.

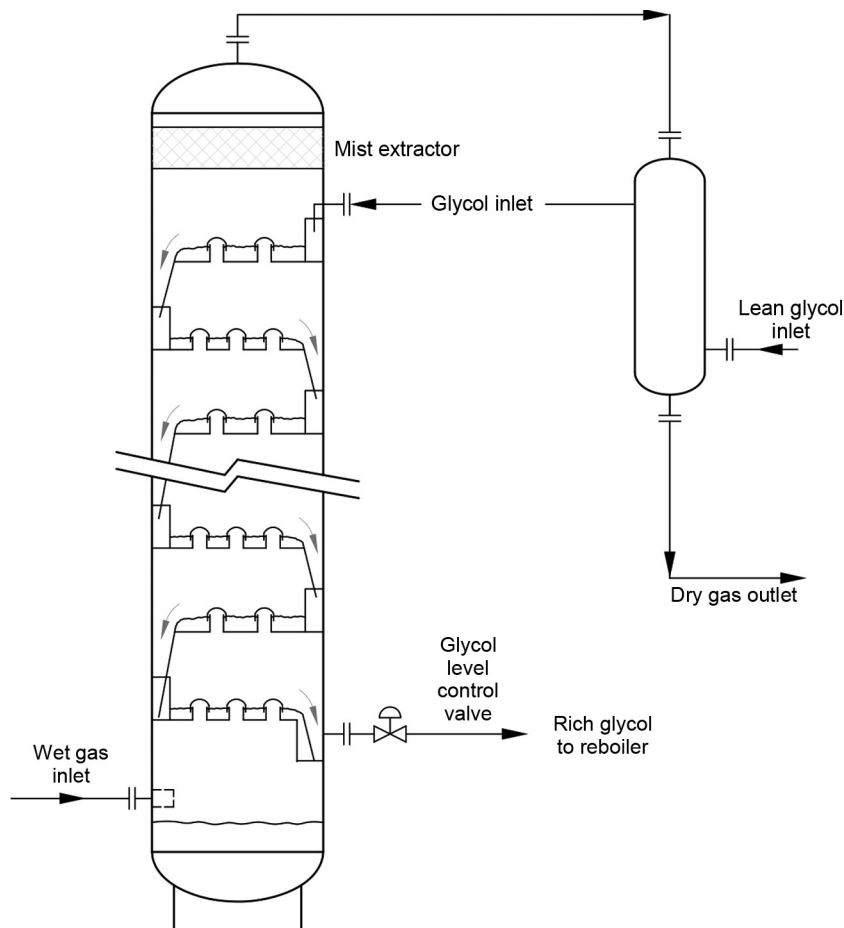


Figure 7.14 Gas system.

7.4.2.2 Glycol Gas Contactor

After passing through the microfiber filter separator, the gas enters the glycol gas contactor near the bottom of the vessel. The inside of the contactor contains either packing or several trays with weirs that maintain a specific level of glycol so that the gas must bubble through the glycol as the gas flows up (Figure 7.16). As the wet gas passes upward through each succeeding tray, it releases the water vapor to the glycol and becomes progressively drier. Before leaving the contactor the gas passes through a mist extractor to remove glycol that may be trying to leave the gas. Dry gas exits the contactor

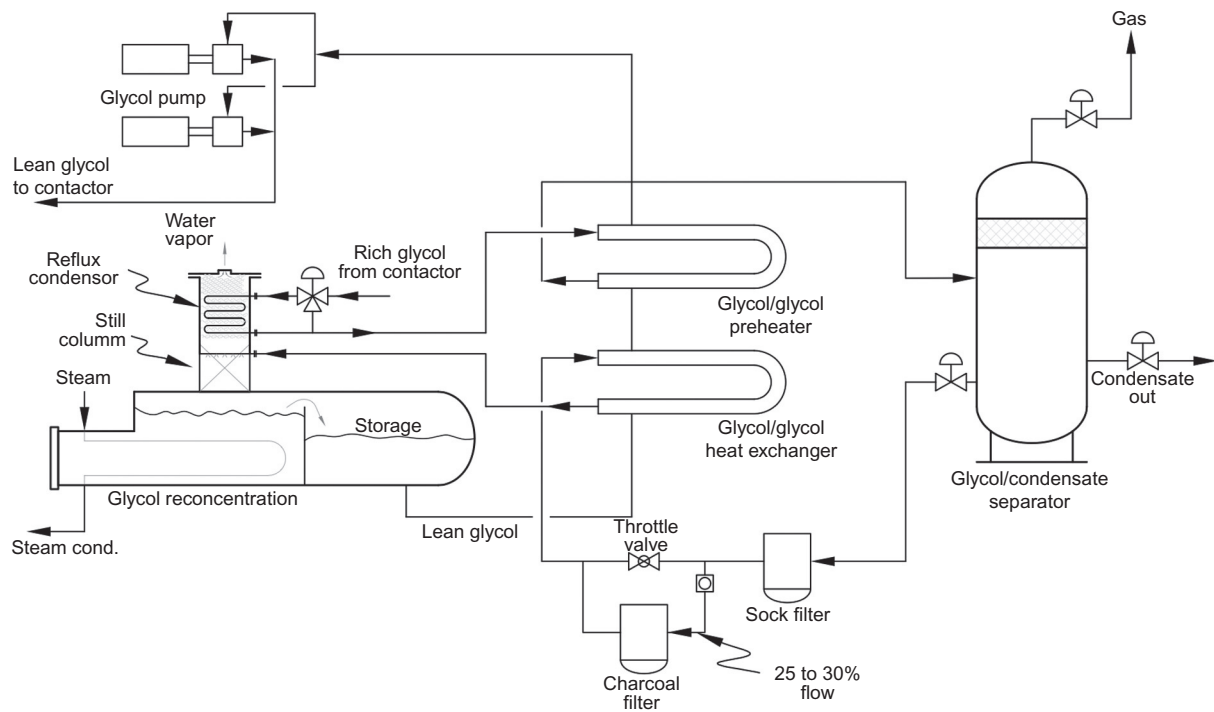


Figure 7.15 Glycol system.

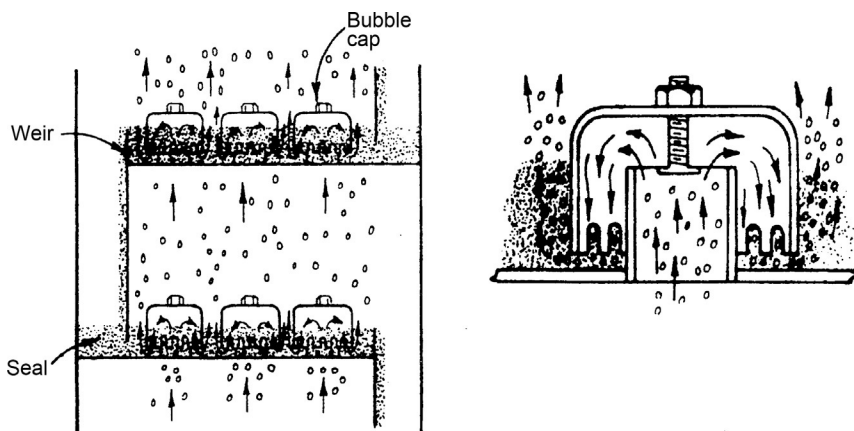


Figure 7.16 Bubble cap trays.

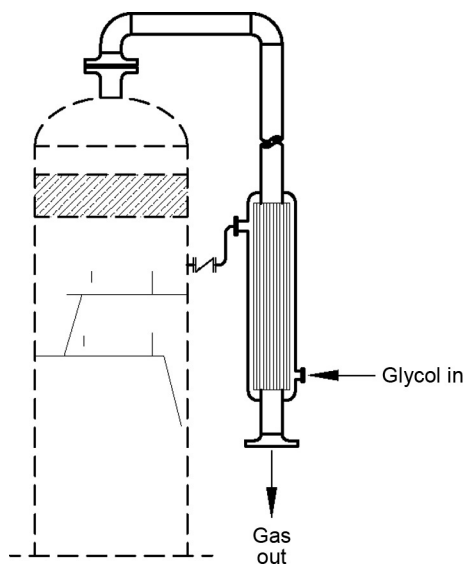


Figure 7.17 External glycol gas heat exchanger.

at the top and passes through an external glycol gas heat exchanger where it cools the incoming dry glycol to increase its absorption capacity (Figure 7.17).

Some installations incorporate a glycol knockout drum (centrifugal separator), which recovers any glycol that has escaped with the gas through the mist extractor (Figure 7.18). The dry gas then leaves the dehydrator unit.

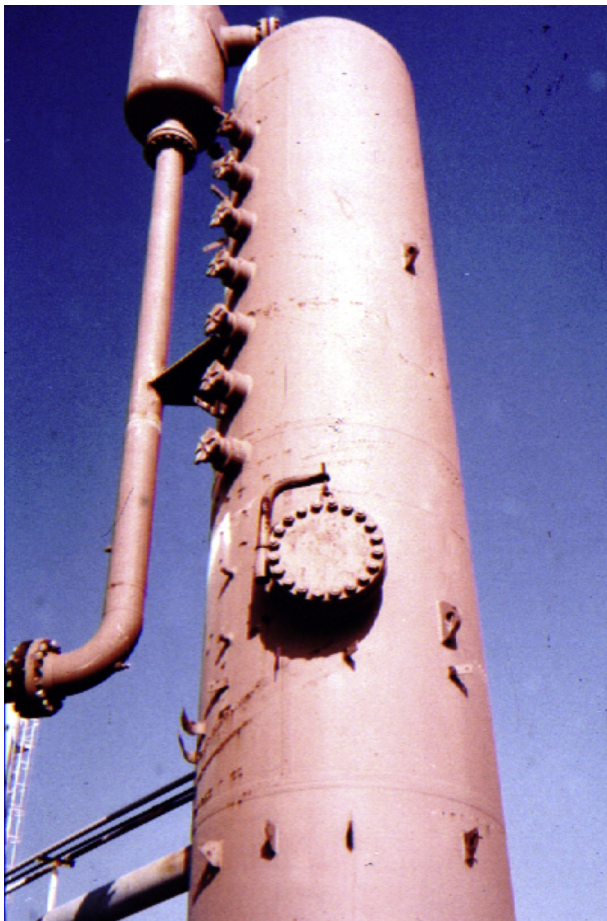


Figure 7.18 Centrifugal separator used to recover escaping glycol with gas.

7.4.3 Glycol System

7.4.3.1 Glycol Gas Heat Exchanger

Dry concentrated glycol is pumped up to contactor pressure, by the glycol pump, and then passes through the glycol gas heat exchanger before entering the contractor tower. The glycol gas heat exchanger cools the glycol to near the temperature of the gas before the glycol enters the contactor. It is important that the glycol be near the gas temperature to prevent gas from exceeding equilibrium temperature and to prevent foaming.

7.4.3.2 Glycol Gas Contactor

Dry glycol from the glycol gas heat exchanger enters the contactor tower and flows across the top tray. This is the first contact between the glycol and the gas. Glycol flows downward through downcomers in the tower, absorbing more water as it passes across each tray. The downcomer seals the glycol passage into the tray below, thus preventing gas from short-circuiting past the bubble caps.

As the glycol flows downward through each succeeding tray, it becomes wetter with the water it has absorbed from the gas and collects in the bottom of the contactor saturated with water.

As the gas moves upward through each succeeding tray, it becomes drier. The wet gas that has accumulated in the bottom of the contactor passes through a strainer (filter), which removes abrasive particles, before flowing through the power side of the glycol pump (energy exchange pumps), where it furnishes the power to pump the dry glycol into the contactor.

Power comes from the increased head caused by the absorbed gas contained in the rich glycol.

7.4.3.3 Reflux Condenser

From the glycol gas contactor, the cool wet glycol passes through a coil (reflux condenser) in the top of the reboiler still column. The coil cools the vapors leaving the still column and condenses the glycol vapors to liquid. The glycol liquid droplets gravitate back down the still column to the reconcentrator. The water remains as a vapor and continues out the top of the still column. The cooling coil is commonly called the reflux condenser.

7.4.3.4 Glycol-Glycol Preheater

The slightly warmed wet glycol leaving the reflux condenser passes through the glycol-glycol preheater. The hot dry glycol from the glycol reconcentrator heats the wet glycol further and, in turn, further cools the dry glycol before it goes to the glycol pumps.

7.4.3.4.1 Gas-Glycol-Condensate Separator

After leaving the glycol-glycol preheater, the heated wet glycol is sent to a low-pressure gas-glycol-condensate separator where most of the entrained gas and liquid hydrocarbons that were picked up by the glycol on its path through the contactor are removed. The heat provided by the glycol-glycol preheater helps in the separation of hydrocarbons from the wet glycol.

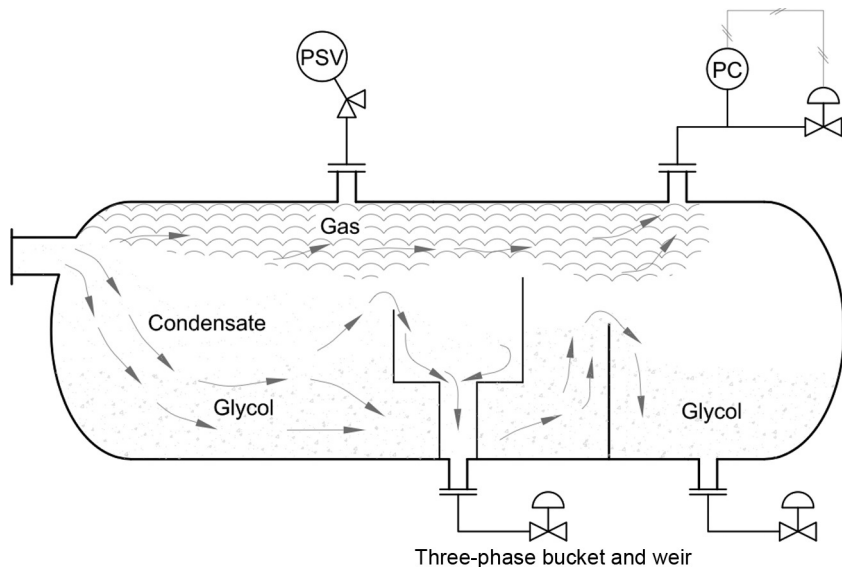


Figure 7.19 Gas-glycol-condensate separator.

The hydrocarbon condensate is separated from the glycol by a three-phase gas-glycol-condensate separator (Figure 7.19).

7.4.3.5 Microfiber Filter

After the gas and condensate has been separated in the gas-glycol-condensate separator, the wet glycol passes through a microfiber filter (Figure 7.20).

7.4.3.6 Charcoal (Carbon) Filter

From the microfiber filter, the wet glycol enters a charcoal, or carbon, filter. Activated carbon granules in this filter absorb liquid-entrained hydrocarbons, well-treating chemicals, compressor oils, and other impurities that may cause foaming.

7.4.3.7 Glycol-Glycol Heat Exchanger

From the charcoal filter, the wet glycol flows through the dry glycol to the wet glycol heat exchanger. This heat exchanger preheats the wet glycol as much as possible before entering the glycol reconcentrator, thus reducing the heat duty of the glycol reconcentrator.



Figure 7.20 Microfiber filters used to remove solids, tarry hydrocarbons, or other impurities.

7.4.3.8 Still Column

From the glycol/glycol heat exchanger, the wet glycol enters the still column that sits vertically on top of the glycol reconcentrator ([Figure 7.21](#)). The inside of the still column is packed with either ceramic saddles or stainless steel pall rings, which are used to add surface area and distribute heat to the incoming glycol. The incoming wet glycol spreads out uniformly and drips down through the packed section. The vapors traveling upward from the glycol reconcentrator heat the packing. As the glycol travels down through the heated packing, water begins to be driven off as steam.

Units utilizing efficient heat exchangers may remove as much as 75–80% of the water contained in the glycol in the still column before the glycol reaches the reconcentrator. As water vapor travels up through the still column and exits from the top, it carries with it trapped glycol vapor. To prevent the loss of glycol vapor, the still column utilizes a “reflux condenser” located on top of the packed still column. Glycol vapors escaping the still column with the steam are attracted to the film of condensed liquid (primarily water) covering the coil surface area where they too are condensed. The liquid droplets gravitate back down the still column into the reconcentrator for further treating, thus preventing excessive glycol loss due to vaporization.

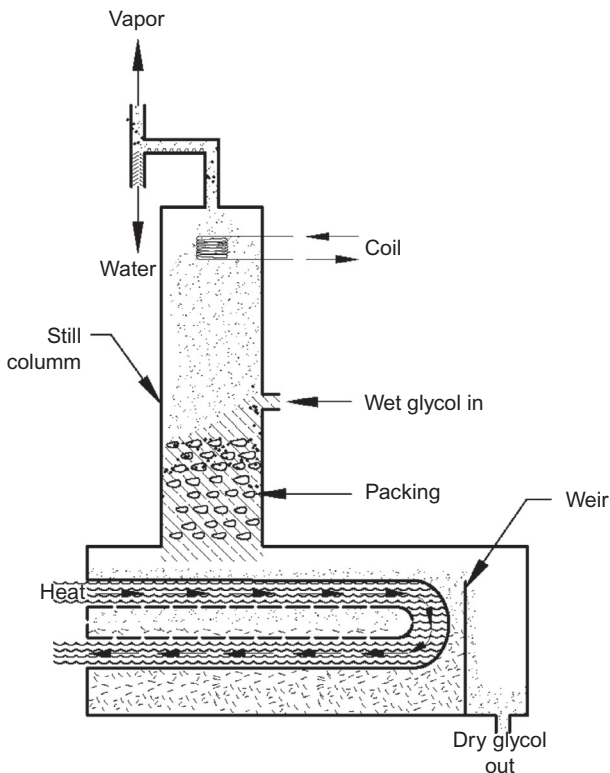


Figure 7.21 Still column at the top of a glycol reconcentrator.

On some units, the glycol enters the still column below the packed section of the column. Vaporization takes place in the reconcentrator. The reflux condenser operates the same in both types of still columns. The packed section is no longer used to distribute heat for vaporization. Condensed liquid from the reflux condenser drops back into the packed section providing a liquid film over the upper portion of the packing. Glycol vapors escaping with steam from the reconcentrator must pass through the packed section. The watery film covering the packing recaptures the glycol vapor, condensing it into droplets, which wash back into the reconcentrator. Thus, more glycol vapor can be recovered in this configuration than in the previously described still column. Since vaporization occurs primarily in the reconcentrator, the operating temperature is lower in this type of still column. This translates into greater reflux condensation and requires larger heat duty.

7.4.3.9 Reconcentrator

From the packed still column, the wet glycol drops downward into the reconcentrator. The glycol is heated to a temperature at which most of the remaining water and some of the glycol are vaporized. A heat source heats the glycol to between 350 and 400 °F, which removes the remaining water and ensures the temperature is below the decomposition point of TEG. The temperature of the glycol in the reconcentrator is critical and must be controlled at this point.

Sources of heat include direct fired (natural draft/forced draft), waste heat (exhaust gases from turbines, engines, and/or generators), steam, and electric heaters. The heated vapor (both glycol and water) rises upward through the still column. As the mixture passes the cool reflux condenser coils, the glycol vapors are condensed and drop back down. The water vapor leaves the top of the still column as steam. Some of the steam will condense, so a downspout is provided to drain the water off.

A weir maintains a level of glycol over the heat source, which prevents overheating of the tubes and prevents premature tube failure. As the glycol is purified, it spills over the weir into a separate compartment. From the reconcentrator, the dry (lean) glycol flows to the accumulator surge tank when the glycol pump raises it to contactor pressure to start another cycle.

7.4.3.10 Stripping Gas

Purities of 98% or more are normally achieved in a TEG system operating at atmospheric pressure. If very pure glycol (up to 99.9% TEG) is required and cannot be achieved by the standard regeneration system, stripping gas may be used. A small amount of dry natural gas, normally taken from the fuel stream, is injected into the reconcentrator. Because hot gas has an affinity for water, the stripping gas is bubbled through the hot glycol, which strips the remaining water from the glycol. This gas can be put directly into the reconcentrator or it can be added to the storage tank where it can percolate through the packed column between the two vessels (Stahl column). The Stahl column also serves as a weir where the dry glycol spills downward by gravity over packing while the gas goes upward, removing even more water. This method prevents air from coming into contact with the dry glycol in the storage tank, thus preventing oxidization of the glycol.

Oxygen entry into the glycol system will decompose the glycol to some extent and cause corrosion within the system. Stripping gas can reduce the temperature at which the reconcentrator must operate and reduce the glycol circulation rate necessary to dehydrate the gas adequately.

7.4.4 Effect of Operating Variables

7.4.4.1 General Considerations

Several operating and design variables have an important effect on the successful operation of a glycol dehydration system.

7.4.4.2 Glycol Selection

Glycols are the most commonly used liquid desiccants in the absorption process because they are:

- Highly hygroscopic (readily absorb and retain water)
- Stable to heat and provide chemical decomposition at the temperature and pressures necessary in the process
- Low vapor pressures, which minimize equilibrium loss of the glycol in the residual natural gas stream and in the regeneration system
- Easily regenerated (water removed) for reuse
- Noncorrosive and nonfoaming at normal conditions (impurities in the gas stream can change this, but even then inhibitors can help to minimize these problems)
- Readily available at moderate cost

Hygroscopicity of glycols is affected by the concentration, or glycol-to-water ratio, that is, increasing as the concentration increases. Dew-point depression obtainable in a gas stream increases as the glycol concentration increases.

7.4.4.2.1 Ethylene Glycol

Ethylene glycol (EG) tends to have high vapor losses to gas when used in a contactor. It is used as a hydrate inhibitor where it can be recovered from the gas by separation at temperatures below 50 °F.

7.4.4.2.2 Diethylene Glycol

Diethylene glycol (DEG) reconcentrates at temperatures between 315 and 325 °F, which yields purity of 97.0%. It degrades at 328 °F. It cannot achieve the concentration required for most applications.

7.4.4.2.3 Triethylene Glycol

TEG is most commonly used in glycol dehydration. It reconcentrates at temperatures between 350 and 400 °F, which yields purity of 98.8%. It degrades at 404 °F and tends to experience high vapor losses to gas at temperatures in excess of 120 °F. With stripping gas, dew-point depressions up to 150 °F are possible.

7.4.4.2.4 Tetraethylene Glycol

Tetraethylene glycol (TTEG) is expensive. It reconcentrates at temperatures between 400 and 430 °F. It experiences lower vapor losses to gas at high gas contactor temperatures, and it degrades at 460 °F.

7.4.5 Inlet Gas Temperature

At constant pressure, the water content of the inlet gas increases as the temperature increases. For example, at 1000 psia and 80 °F, gas holds 34 lbs. of water/MMSCF and at 120 °F, gas holds 104 lbs. of water/MMSCF.

If the gas is saturated at the higher temperature, the glycol will have to remove about three times as much water to meet the specifications. Temperatures above 115 °F result in high glycol losses and thus require TTEG. Temperature should not fall below the hydrate formation temperature range (65–70 °F) and should always be above 50 °F. Temperatures below 50 °F cause problems due to an increase in glycol viscosity. Temperatures below 60–70 °F can cause a stable emulsion with liquid hydrocarbons in the gas and cause foaming in the contactor. An increase in gas temperature increases the gas volume, which in turn increases the diameter of the glycol contactor.

7.4.6 Lean Glycol Temperature

Dry glycol temperatures entering the top tray of the contactor (approach temperature) should be held low (10–15 °F) above the inlet gas temperature. Equilibrium conditions between the glycol and the water vapor in the gas is affected by temperature. Glycol entering the top tray of the contactor may raise the temperature of the gas surrounding it and prevent the gas from releasing its remaining water vapor. Inlet glycol temperatures >15 °F above the gas temperature results in high glycol losses to the gas. Drastic temperature differential also has a tendency to emulsify the glycol with any contaminants resulting in subsequent glycol loss.

7.4.7 Glycol Reconcentrator Temperature

Reconcentrator temperature controls the concentration of the water in the glycol. With a constant pressure, the glycol concentration increases with higher reconcentrator temperature. Reconcentrator temperature should be limited to between 350 and 400 °F because it minimizes degradation of TEG, which begins to degrade at 404 °F, and it results in lean glycol concentrations between 98.5% and 98.9%.

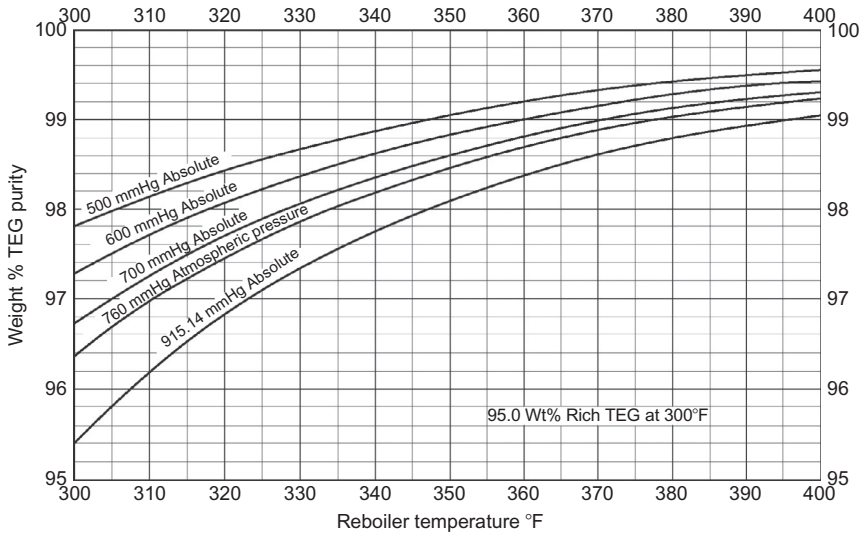


Figure 7.22 Glycol purity versus reconcentrator temperature at different levels of vacuum.

Figure 7.22 shows the glycol concentrations that can be obtained with various reboiler temperatures. When higher lean glycol concentrations are required, stripping gas should be added to the reconcentrator and/or the reconcentrator and still column should be operated in a vacuum.

7.4.8 Temperature at the Top of the Still Column

A high temperature in the top of the still column can increase glycol losses due to excessive vaporization. A reboiler temperature in the range of 350–400 °F ensures adequate heat transfer to the ceramic packing in the still column. The still column operates best (allows the steam to escape) when the vapor outlet temperature is between 215 and 225 °F. When the temperature reaches 250 °F and above, glycol vaporization losses increase. The still top temperature can be lowered by increasing the amount of glycol flowing through the reflux condenser coil.

If the temperature in the top of the still column drops too low (below 220 °F), too much water can be condensed and washed back into the reconcentrator, which increases the reconcentrator heat duty. Too much cool glycol circulation in the reflux condenser coil can lower the still top temperature below 220 °F, which can cause the excess water to condense. Thus, most reflux condenser coils have a bypass valve, which allows manual or automatic control of the stripping still temperature.

7.4.9 Contactor Pressure

At a constant temperature, the water content of the inlet gas decreases with an increase in pressure. The lower the pressure, the larger the contactor diameter required. Good dehydration can be achieved at any pressure below 3000 psig as long as the pressure is constant. Optimum dehydration pressure is often in the range of 550–1200 psig. Sizing calculations should always be based on minimum expected operating gas pressure.

Rapid pressure changes translate into rapid velocity changes in the contactor, which breaks the liquid seals between the downcomers and the trays, allows the gas to channel up through both the downcomer and bubble caps, and allows the glycol to be swept out with the gas.

7.4.10 Reconcetrator Pressure

Reducing the pressure in the reconcentrator at a constant temperature results in higher glycol purity. Most reconcentrators operate between 4 and 12 ounces of pressure. On standard atmospheric reconcentrators, pressures in excess of 1 psi result in glycol loss from the still column, reduction of lean glycol concentration, and reduction in dehydration efficiency.

Pressures more than 1 psi are usually associated with excess water in the glycol and create a vapor velocity exiting the still great enough to sweep glycol out. Fouled still column packing often contributes to high reconcentrator pressure. The still column should be adequately vented and the packing replaced periodically so as to prevent back pressure on the reconcentrator.

Pressures below atmospheric will increase the lean glycol concentration because the boiling temperature of the rich glycol/water mixture decreases. Reconcetrators are rarely operated in a vacuum due to the added complexity and the fact that air leaks will result in glycol degradation.

7.4.11 Contractor Pressure

If lean glycol concentrations in the range of 99.5% are required consider operating the reconcentrator at a pressure 500 mm Hg absolute (10 psia) or using stripping gas.

Figure 7.22 can be used to estimate the effect of operating in a vacuum on lean glycol concentration.

7.4.12 Glycol Concentration

The water content of the dehydrated gas depends primarily on the lean glycol concentration. The higher the concentration of lean glycol entering the

contactor, the greater the dew-point depression for a given circulation rate and number of trays.

Increasing the glycol concentration above 99% purity can lead to dramatic results on the outlet dew point (Figure 7.23). For example, with a 100 °F inlet gas temperature (110 °F top tray temperature), an outlet dew point of 10 °F can be obtained with 99.0% TEG, -30 °F can be obtained with 99.8% TEG, and -40 °F can be obtained with 99.9% TEG.

Higher concentrations of TEG can be obtained by increasing the glycol reconcentration temperature, injecting stripping gas into the reconcentrator, or reducing the operating pressure of the reconcentrator.

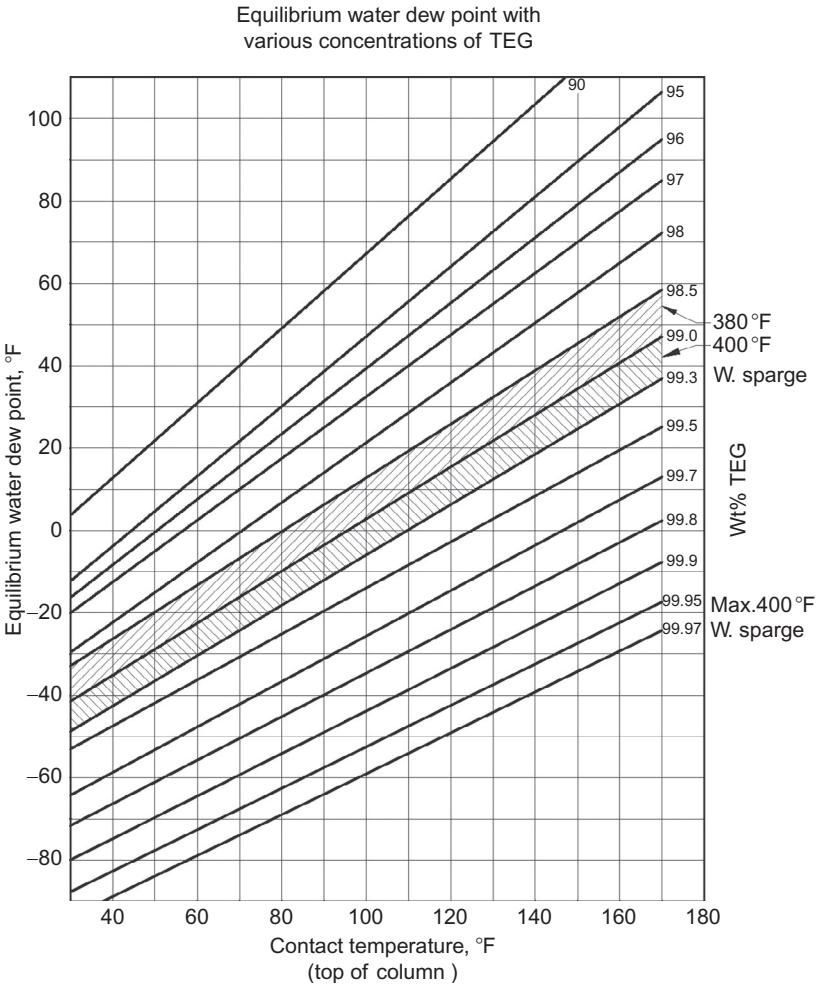


Figure 7.23 Equilibrium water dew points with various concentrations of TEG.

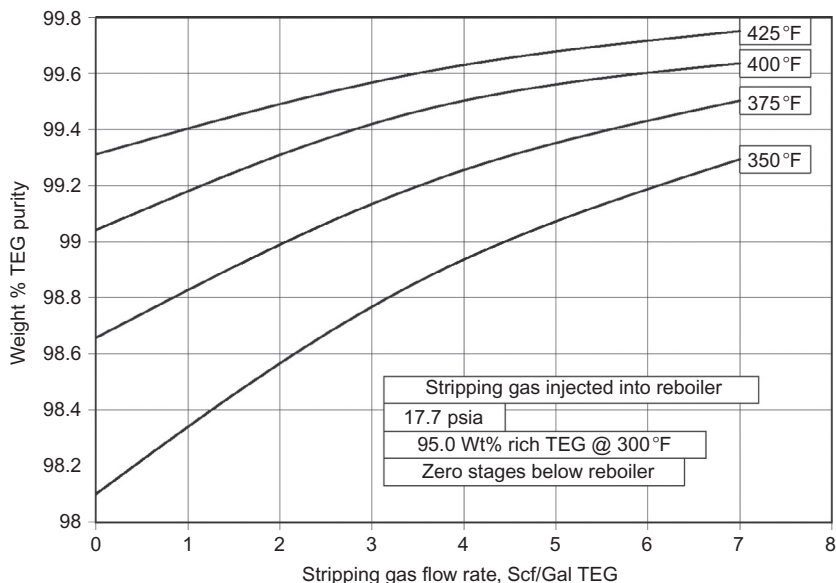


Figure 7.24 Effect of stripping gas on TEG concentration.

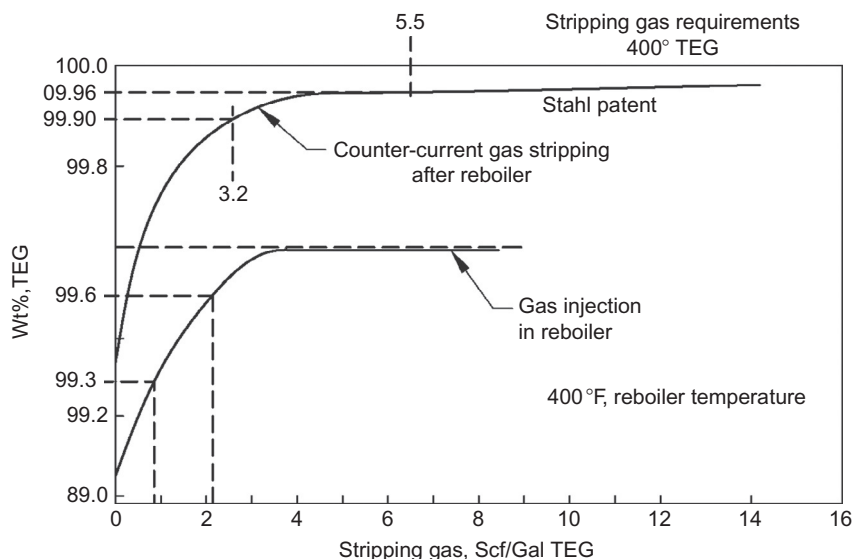


Figure 7.25 Effect of stripping gas on the concentration using the Stahl column.

Reconcentration temperatures for TEG normally run between 380 and 400 °F, which results in glycol purities of 98–99%. [Figures 7.24](#) and [7.25](#) illustrate the effect of stripping gas. If gas is injected directly into the reconcentrator (via a sparger tube), the concentration of TEG increases

significantly from 99.1% to near 99.6% as the gas rate is increased from 0 to 4 SCF/gal.

When the Stahl method is used (counter-current gas stripping after the reconcentrator), concentrations as high as 99.95% TEG can be attained at a 400 °F reconcentrator temperature.

7.4.13 Glycol Circulation Rate

When the number of absorber trays and the lean glycol concentration are held constant, the dew-point depression of a saturated gas is increased as the glycol circulation rate is increased. The more lean glycol that comes into contact with the gas, the more water vapor is absorbed out of the gas.

Whereas the glycol concentration mainly affects the dew point of the dry gas, the glycol rate controls the total amount of water that can be removed. The normal operating level in a standard dehydrator is three gallons of glycol per pound of water removed (range two to seven).

Figure 7.26 shows that a greater dew-point depression is easier to achieve by increasing the glycol concentration rate. Excessive circulation rates:

- Overload the reconcentrator
- Prevent good glycol regeneration

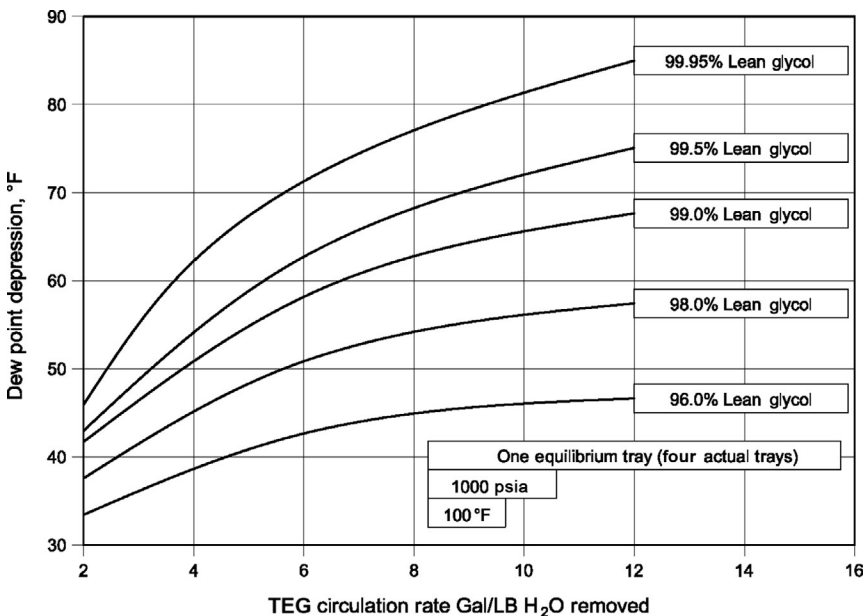


Figure 7.26 Calculated dew-point depression versus circulation rate (1 equilibrium tray (4 actual trays)).

- Prevent an adequate glycol gas contact in the absorber
- Increase pump maintenance problems
- Increase glycol losses

Heat required by the reboiler is directly proportional to the circulation rate. An increase in circulation rate may decrease the reconcentrator temperature, decrease the lean glycol concentration, or decrease the amount of water that is removed by the glycol from the gas. Only if the reconcentrator temperature remains constant will an increase in circulation rate lower the dew point of the gas.

7.4.14 Number of Absorber Trays

When the glycol circulation rate and the lean glycol concentration are held constant, the dew-point depression of a saturated gas is increased as the number of trays is increased. Actual trays do not reach equilibrium, and their approach to it is expressed as a fraction of a theoretical tray. A tray efficiency of 25% is commonly used for design. Four actual trays with efficiencies of 25% would accomplish the job of one theoretical tray. The number of actual trays in a design ranges from 4 to 12. An approximation of the actual number of valve trays per foot of packing can be obtained from [Figure 7.27](#).

For high-performance units, the specification of more than four trays in a new design can achieve fuel savings (for the same dew-point depression) due to lower circulation rate, lower reconcentration temperature, or lower stripping gas rate.

[Figure 7.28](#) shows that specifying a few additional trays in the contactor is much more effective than increasing the glycol circulation rate. The additional investment for a taller contactor is often justified by fuel savings.



7.5 SYSTEM DESIGN

7.5.1 Sizing Considerations

Sizing involves specifying the following:

- Glycol gas contactor diameter
- Number of absorber trays (establishes the tower's overall height)
- Glycol circulation rate
- Lean glycol concentration
- Reconcentrator heat duty

The number of absorber trays, glycol circulation rate, and lean glycol concentration are all interrelated.

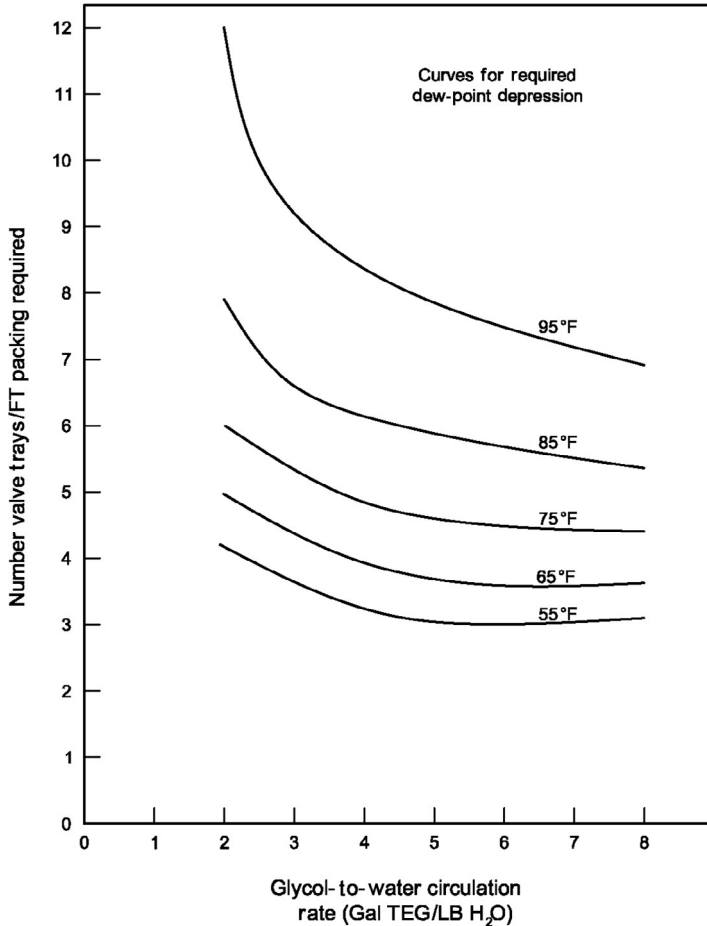


Figure 7.27 Trays of packing required for glycol dehydrator.

7.5.2 Inlet Microfiber Filter Separator

Clean inlet gas is a key factor in minimizing absorber operating problems. By eliminating liquid water carryover, the inlet scrubber (microfiber filter separator) can prevent dilution of the glycol, lower absorber efficiencies, high glycol circulation rates, high vapor-liquid loads on the still column, still column flooding, and high reboiler heat-load and fuel-gas requirements. These problems also cause high glycol losses and wet sales gases. The scrubber (microfiber separator) also prevents salt or other solids from entering the glycol system where they could be deposited in the reconcentrator to foul the heating surfaces and burn out as hot spots. The scrubber should be sized

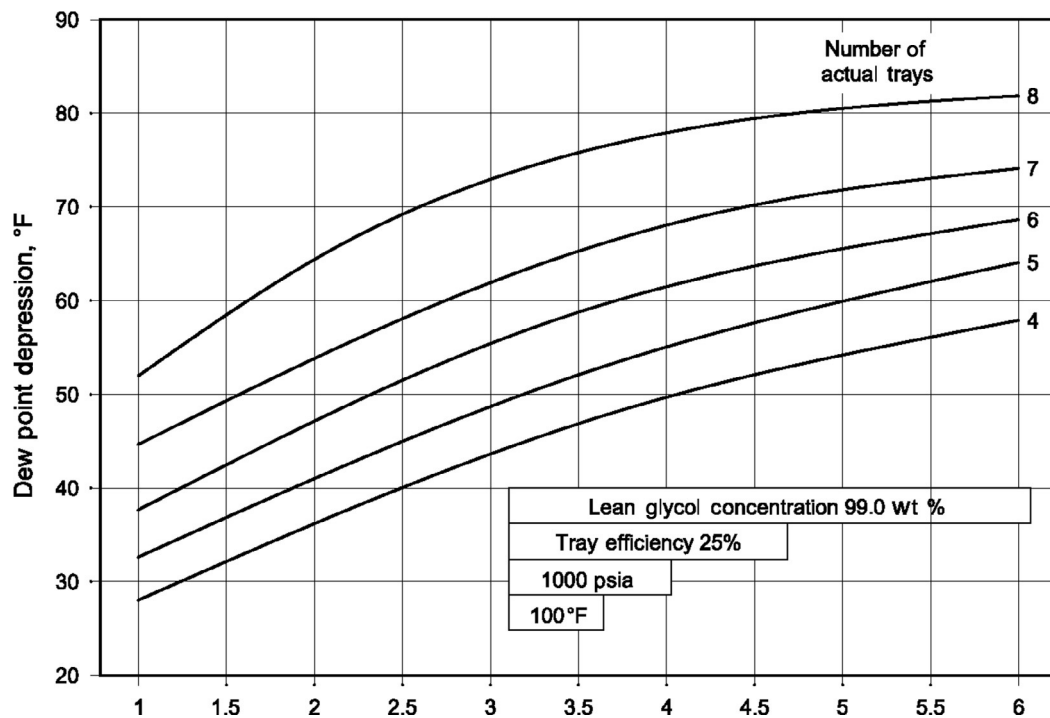


Figure 7.28 Effect of number of absorber trays on dew-point depression.

according to procedures covered in [Chapter 4](#) of Volume 1 for gas-liquid separation sizing.

The vessel cleans the incoming gas and is useful when paraffin and other impurities are present in a fine vapor form. A mist extractor, sized to remove 99% of all contaminants over one micron in size, is used to clean the incoming gas.

A microfiber filter separator is highly recommended when inlet gas streams have been compressed or if structured packing is used in the contactor. Compressor oil and heavy distillate can coat the tower packing either in the contactor or the still column, thus decreasing the effectiveness of the equipment.

7.5.3 Glycol Gas Contactor

The two basic types of glycol gas contactor are trayed towers and packed towers.

TEG is viscous (resulting in poor tray efficiency) and exhibits foaming tendencies (restricts tower performance). Due to the low liquid loadings, it is usually possible to provide a high downcomer residence time with only a small percentage of the tower area dedicated to downcomers.

Some contactors have an “internal scrubber,” which occupies approximately the lower one-third of the vessel. They are usually installed on units where the inlet gas flow rate is <50 MMSCFD.

A “Chimney” is included on the scrubber/contactor combination ([Figure 7.29](#)). The chimney consists of a large stack that covers the top of the inlet scrubber. The chimney allows the gas to pass upward from the scrubber section to the absorber section and prevents glycol from being lost out of the scrubber section.

Some contact towers have an internal three-phase separator. These towers are distinguishable in that the lower section has two sets of level controls and two liquid dump valves. This configuration is not recommended as it is difficult to troubleshoot should a problem occur. A separate two-phase microfiber filter separator located immediately upstream of the contactor is the most efficient and trouble-free configuration.

7.5.4 Contactor Diameter

The minimum diameter for trayed towers and conventional packing can be determined from the following equation:

$$d^2 = 5040 \left(\frac{T_o Z Q_g}{P} \right) \left| \left(\frac{\rho_s}{\rho_L - \rho_g} \right) \left(\frac{C_d}{d_m} \right) \right|^{1/2} \quad (7.9)$$

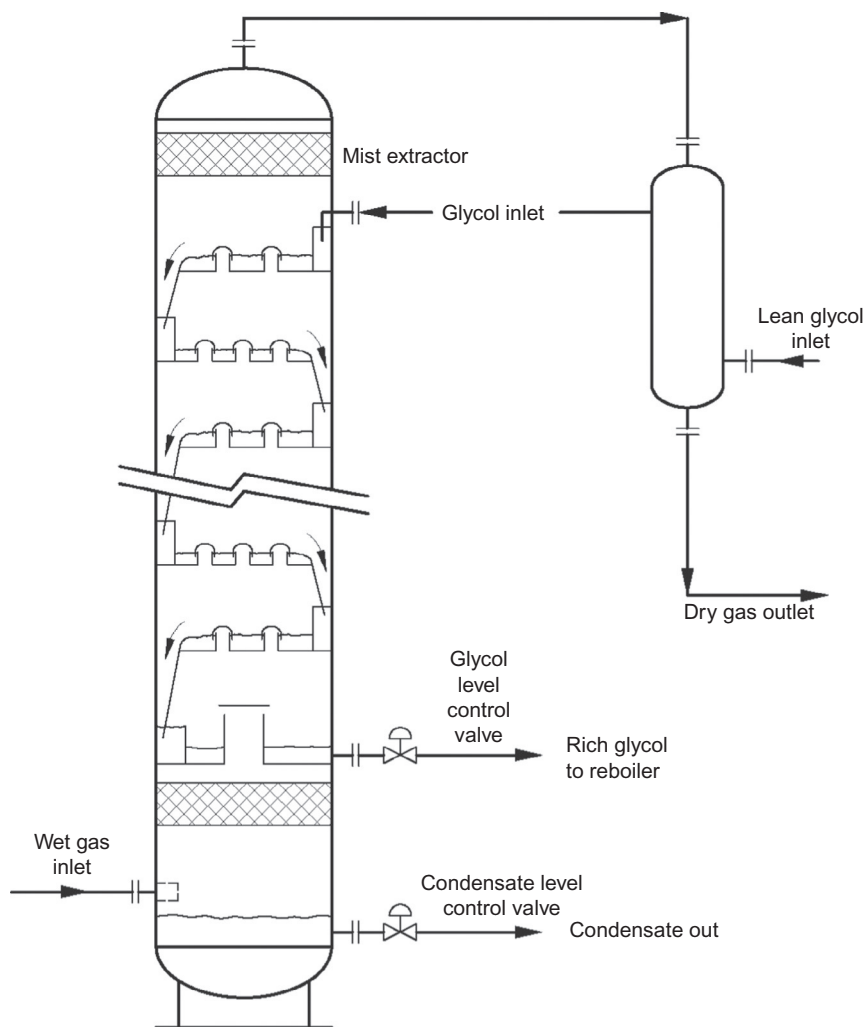


Figure 7.29 Glycol contactor with scrubber section.

where d = contactor inside diameter, inches; d_m = drop size, microns = 120–150 μm range; T_o = contactor operating temperature, $^{\circ}\text{R}$; Q_g = design gas flow rate, MMSCFD; P = contactor operating pressure, psia; C_d = drag coefficient; ρ_g = gas density, $\text{lb}/\text{ft}^3 = 2.7 \text{ (SP/TZ)}$; ρ_L = glycol density, $\text{lb}/\text{ft}^3 = 70 \text{ lb}/\text{ft}^3$; Z = compressibility factor; S = gas specific gravity (air = 1).

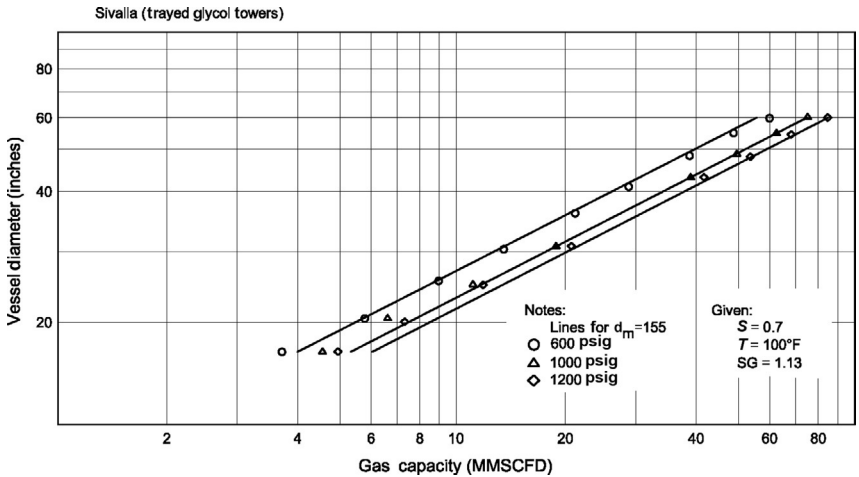


Figure 7.30 Determination of contactor diameter—Sivalls.

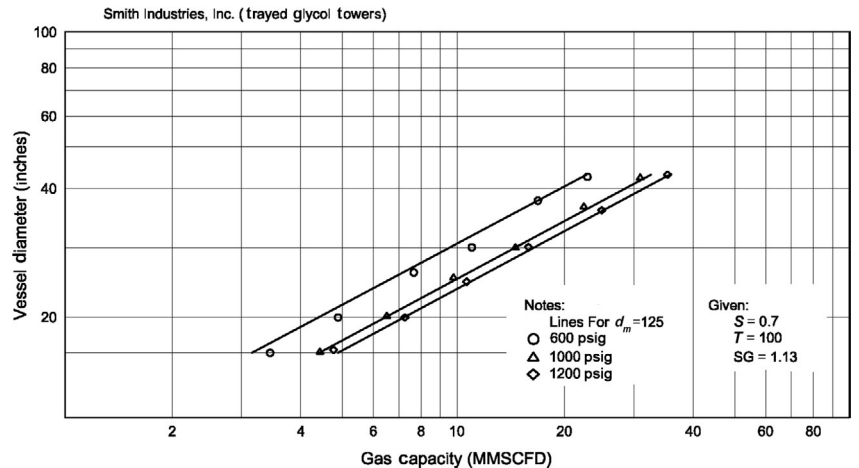


Figure 7.31 Determination of contactor diameter—Smith Industries.

Structured packing can handle higher gas flow rates for the same diameter contactor. Figures 7.30–7.33 are correlations prepared by vessel manufacturers that allow graphical solutions of glycol gas contactor diameters.

7.5.5 Tray Design

7.5.5.1 Bubble Cap Trays (Figures 7.34–7.38)

Bubble cap trays are the most commonly used tray configuration in glycol contactors. They are better than conventional packing (Figure 7.39).

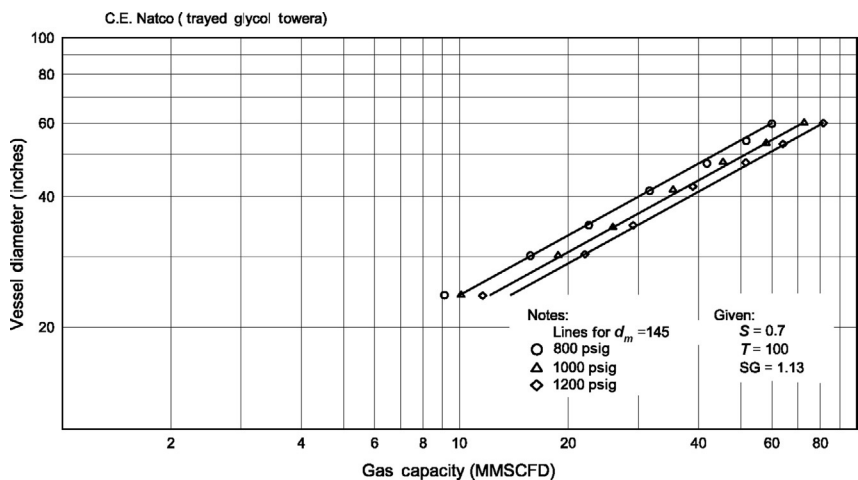


Figure 7.32 Determination of contactor diameter—NATCO.

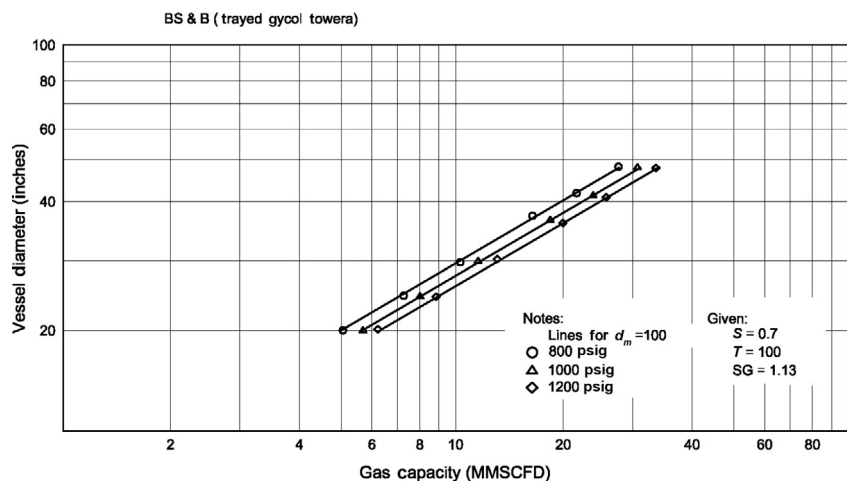


Figure 7.33 Determination of contactor diameter—BS&B.



Figure 7.34 Common bubble cap configurations.



Figure 7.35 Bubble cap components.



Figure 7.36 Bubble cap tray outside the contactor tower.



Figure 7.37 Bubble cap tray inside the contactor tower.



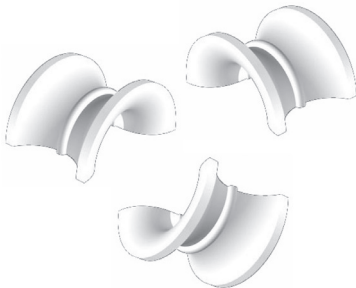
Figure 7.38 Bottom of bubble cap tray.



Raschig ring



Pall ring



Berl saddle

Figure 7.39 Various types of conventional packing.

7.5.5.2 Valve or Flapper Trays (Figures 7.40–7.42)

In the valve, or flapper, tray the gas moves upward through a hole in the bottom of the tray. Over the hole is a device that flutters, or flaps, in an “up and down” manner breaking the gas stream into bubbles that form the froth layer.



Figure 7.40 Top of flapper tray.



Figure 7.41 Bottom of flapper tray.



Figure 7.42 Top and bottom of valve tray.

7.5.5.3 Perforated (Sieve) Trays (Figure 7.43)

The perforated (sieve) tray consists of hundreds of tiny holes. The gas stream passes through these holes and breaks up into the bubbles necessary to form a froth. They are inexpensive to fabricate. The gas capacity range that can be effectively dehydrated is limited.

7.5.5.4 Structured (Matrix) Packing (Figure 7.44 and 7.45)

Structured (matrix) packing resembles corrugated metal affixed side-by-side with the corrugations set at opposite angles. Gas passes upward, through small holes that are drilled in the corrugations, and forms channels by the opposing corrugation. Glycol runs down through the holes and channels contacting the gas. Structured (matrix) packing is the most efficient packing.

7.5.5.5 Tray Spacing

Tray spacing ranges from 20 to 30 in. Preferred spacing is 24 in. but 30-in. spacing is recommended if foaming is anticipated.

7.5.5.6 Number of Trays

Six to eight trays are used to meet normal dew-point depressions. Twelve trays are typically required for high dew-point depressions.

7.5.5.7 Downcomers

Downcomers are sized for a maximum velocity of 0.25 ft./s.

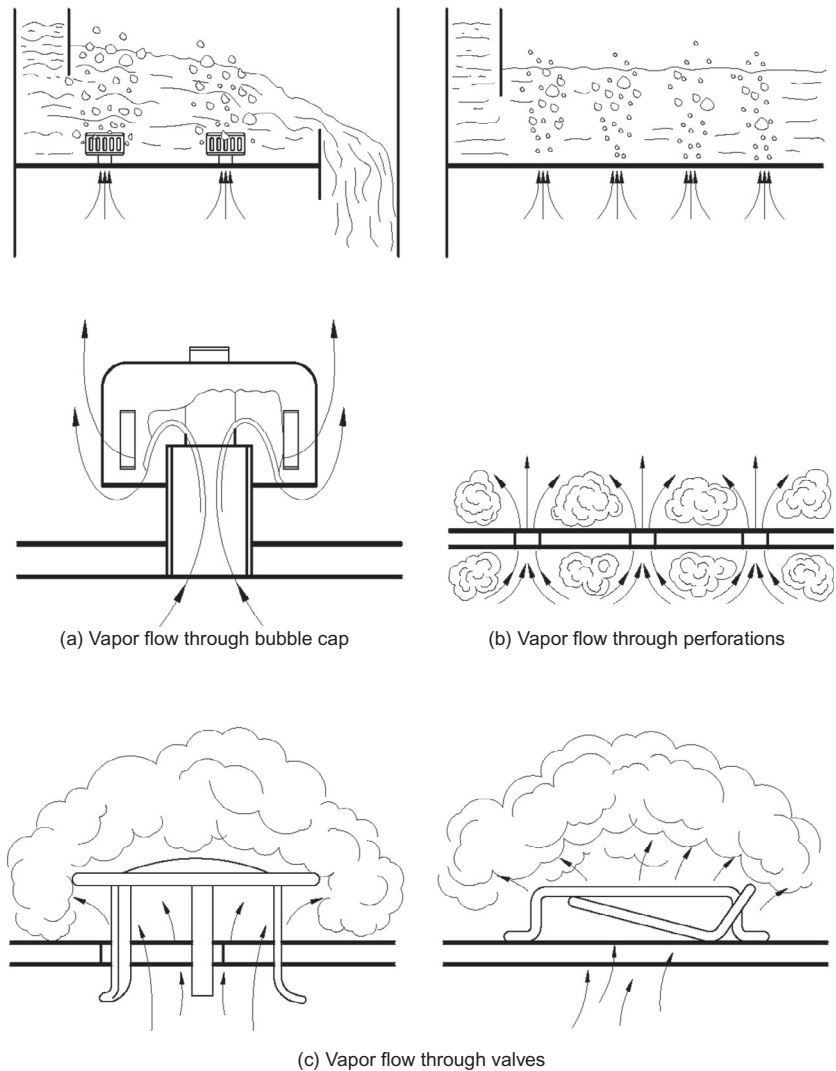


Figure 7.43 Types of valve trays.

7.5.6 Glycol Circulation Rate

For a given dew-point depression, the circulation rate is dependent on the lean glycol concentration and the number of trays. When the lean glycol concentration and the number of trays are held constant, the required glycol circulation rate can be determined from the following equation:

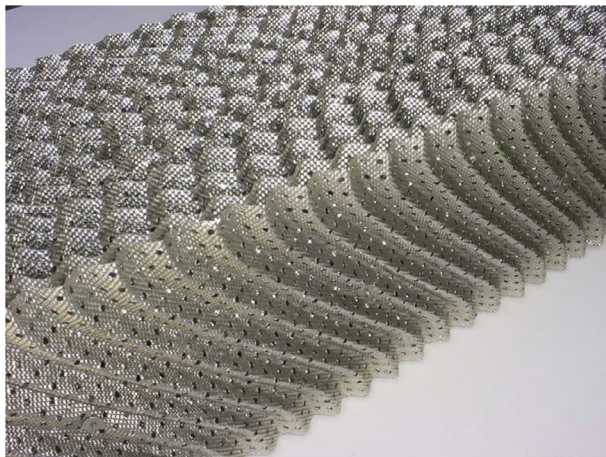


Figure 7.44 Structured packing—side view.

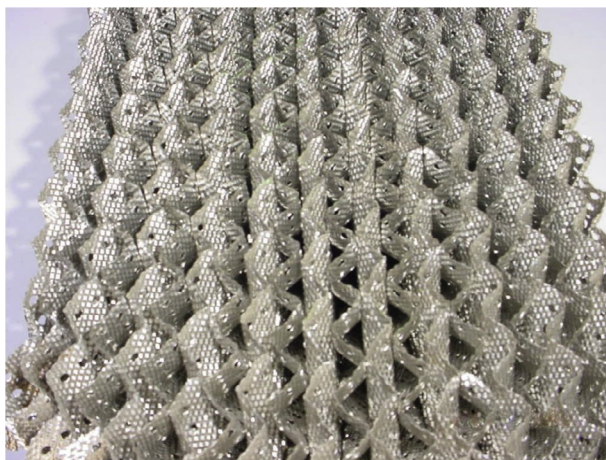


Figure 7.45 Structured packing—top view.

$$L = \frac{\left(\frac{\Delta W}{W_i} \right) W_i Q_g}{24} \quad (7.10)$$

where L = glycol circulation rates, gal/h; $\Delta W/W_i$ = circulation ratio, gal TEG/lb H_2O (see [Figures 7.46–7.48](#)); W_i = water content of inlet gas, lb H_2O /MMSCF; W_o = desired outlet water content, lb H_2O /MMSCF; $\Delta W = W_i - W_o$; Q_g = gas flow rate, MMSCFD.

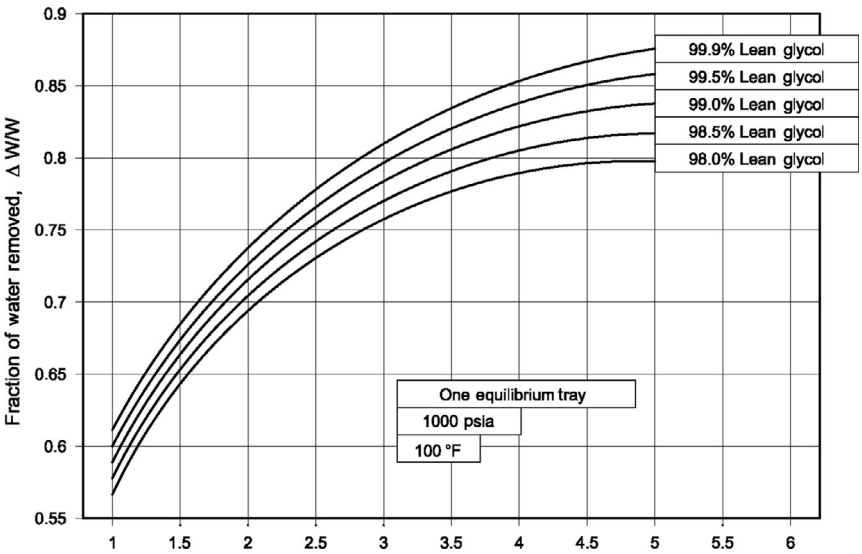


Figure 7.46 Fraction of water removed versus TEG circulation rate ($n = 1$ theoretical trays, four actual trays).

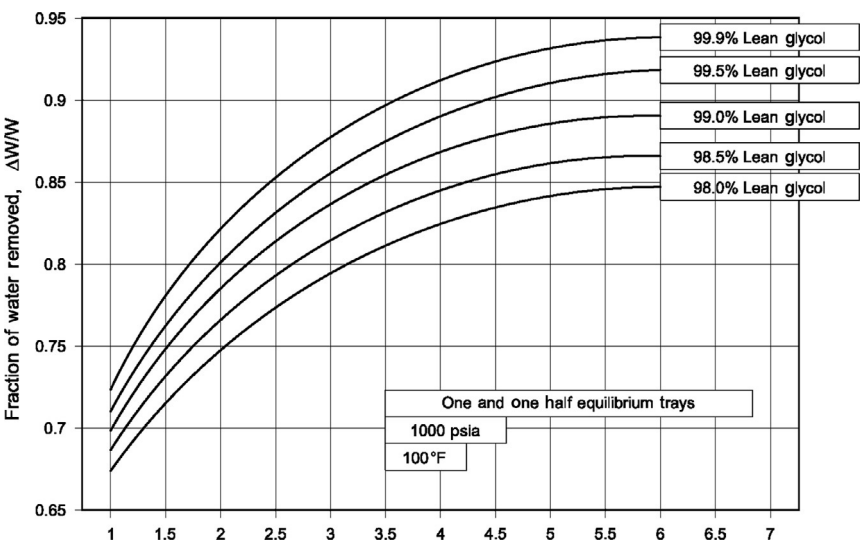


Figure 7.47 Fraction of water removed versus TEG circulation rate ($n = 1\frac{1}{2}$ theoretical trays, six actual trays).

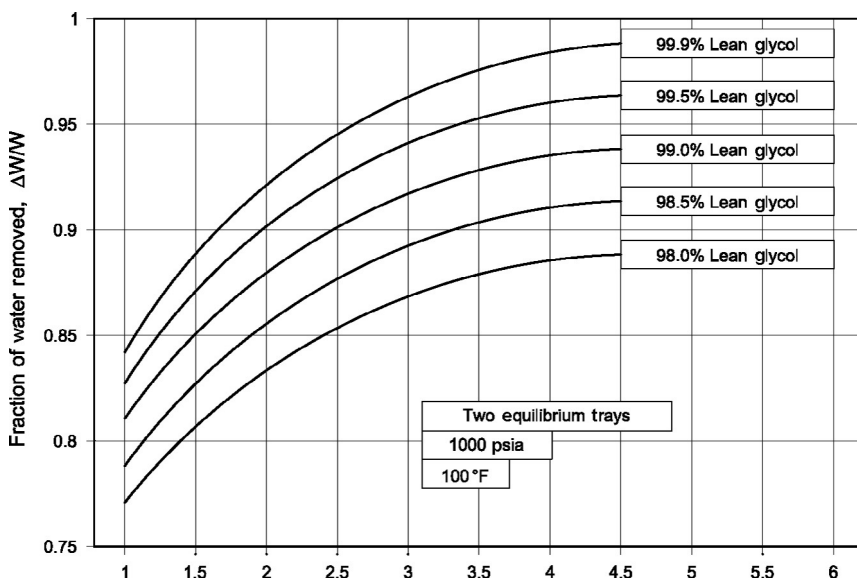


Figure 7.48 Fraction of water removed versus TEG circulation rate ($n=2$ theoretical trays, eight actual trays).

Figures 7.46–7.48 show the fraction of water removed versus TEG rate with respect to different glycol purities.

7.5.7 Lean Glycol Concentration

Equilibrium water dew points for various concentrations of TEG are shown in Figure 7.23. Glycol purity (lean glycol concentration) is a function of the temperature of the reconcentrator (Figure 7.22).

Glycol purity can be increased by adding stripping gas, reducing the pressure in the reconcentrator, and reducing the glycol circulation rate.

7.5.8 Glycol-Glycol Preheater

Cool wet glycol from the contactor enters the preheater (heat exchanger) at 100 °F and the warm glycol leaves at 175–200 °F to the gas-glycol-condensate separator. Hot dry glycol from the glycol/glycol heat exchanger enters the preheater at 250 °F and the warm dry glycol leaves at 150 °F to the glycol pumps on the way to the contactor. Glycol powered pumps (e.g., Kimray) are limited to 200 °F, while electric plunger pumps are limited to 250 °F. The overall heat transfer coefficient is $U=10$ to 12.

7.5.9 Glycol-Gas Cooler

The TEG entering the gas contactor is limited to 10–15 °F above the inlet gas temperature. If hotter, some TEG will vaporize with the gas. If colder, gas condensation of the hydrocarbons may cause foam and glycol loss. The overall heat transfer coefficient is $U=45$.

7.5.10 Glycol-Glycol Heat Exchanger

Hot dry glycol from the reconcentrator enters the heat exchanger at 390 °F and leaves at 250 °F to the glycol/glycol preheater. Warm wet glycol from the charcoal filter enters the heat exchanger at 200 °F and the hot wet glycol leaves at 350 °F to the still column.

7.5.11 Gas-Glycol-Condensate Separator

The separator should be sized using the procedures in [Chapter 4](#) of Volume 1 for sizing gas-liquid separation. Liquid retention times between 20 and 30 min, depending on API gravity of the condensate, are recommended. The operating pressure of 35–50 psig is recommended.

7.5.12 Reconcentrator

The reconcentrator should be designed to operate at 350–400 °F with TEG and 305 °F with DEG. Design temperature should be sufficiently below the decomposition point so that hot spots on the fire tube and poor mixing in the reconcentrator will not cause decomposition of the glycol.

With everything else operating normally, the reconcentrator temperature is raised to lower the water content of the treated gas, and vice versa. Specific reconcentrator operating temperature is determined by trial and error. Temperatures up to 400 °F are common. 400 °F yields 99.5% TEG purity, while 375 °F yields 98.3% TEG purity.

7.5.13 Heat Duty

Heat duty may be estimated using the following equations:

$$q_t = LQ_L \quad (7.11)$$

where q_t = total heat duty on reconcentrator, Btu/h; L = glycol circulation rate, gal/h; Q_L = reconcentrator heat load, Btu/gal TEG.

Heat duty estimated from Equation (7.11) is normally increased by 10–25% to account for start-up, fouling, and increased circulation [Table 7.4](#) lists reconcentrator heat loads.

Table 7.4 Reconcentrator Heat Load
Design (gal TEG/lb H₂O) **Reconcentrator Heat Load
 (Btu/gal TEG circulated)**

2.0	1066
2.5	943
3.0	862
3.5	805
4.0	762
4.5	729
5.0	701
5.5	680
6.0	659

7.5.14 Fire Tube Sizing

The actual surface area of the firebox required for direct-fired heaters can be calculated from the following equation:

$$A = \frac{q_t}{6000} \quad (7.12)$$

where A = total firebox surface area, ft.²; q_t = total heat duty on reconcentrator, Btu/h.

By determining the diameter and overall length of the U-tube fire tube, one can estimate the overall size of the reconcentrator. A heat flux of 6000–8000 Btu/h-ft.² is often used, but the 6000 value is suggested to ensure against glycol decomposition.

7.5.15 Reflux Condenser

Wet glycol inlet from the gas contactor enters at 115 °F and leaves at 125 °F. Controls TEG losses. Reflux rate should be 50% of the water removal rate. The condensing coil allows uniform year round refluxing, provides lowest TEG loss and provides the most economical reconcentrator operation

7.5.16 Stripping the Still Column

Temperature is critical to the operation of the still column. Heat is provided by the reconcentration. Reconcentrator temperatures in the range of 350–400 °F ensures adequate heat transfer to the ceramic packing in the still column.

Still columns in which the wet glycol inlet enters above the packed section (Figure 7.49) operate best with a vapor outlet temperature between 225

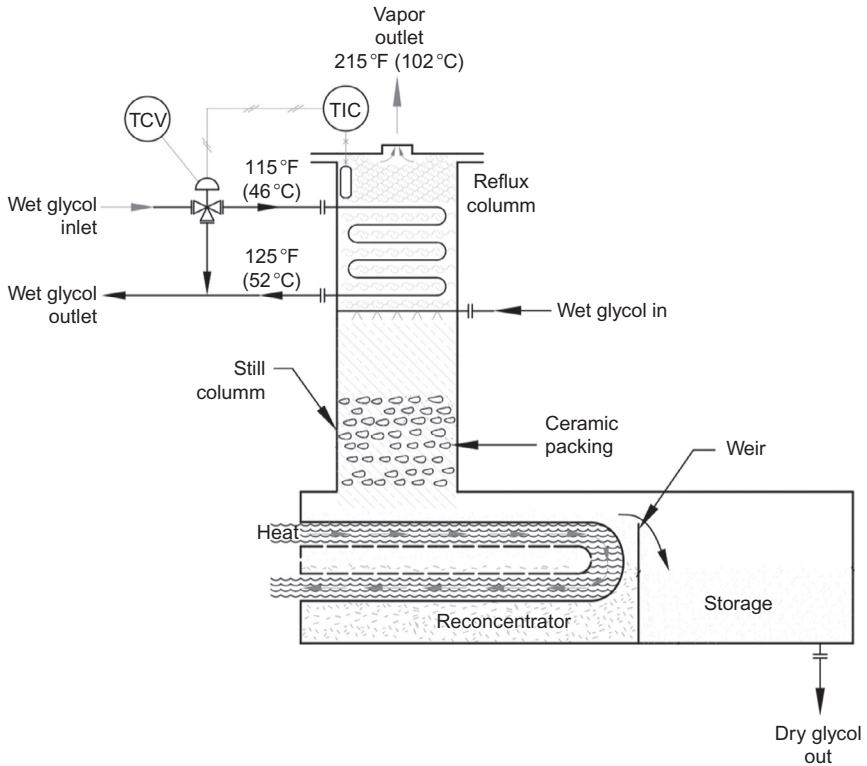


Figure 7.49 Still column with wet glycol entering above the ceramic saddle packing.

and 250 °F. The purpose of glycol falling over ceramic packing is the efficient use of available heat. Backpressure should be kept to a minimum (1 psig is maximum).

Still columns in which the wet glycol inlet enters below the packed section ([Figure 7.50](#)) allow the pall ring type packing to be solely involved in the reflux process. They operate best with a vapor outlet temperature between 185 and 195 °F. This temperature allows a greater volume of condensation by the reflux coil while still permitting the majority of the steam to escape.

7.5.17 Diameter Size

Diameter size is based on the required diameter at the base of the still, calculated by vapor and liquid loading conditions at that point. Vapor load consists of the water vapor (steam) and stripping gas flowing up through the still. Liquid load consists of the rich glycol stream and reflux flowing downward

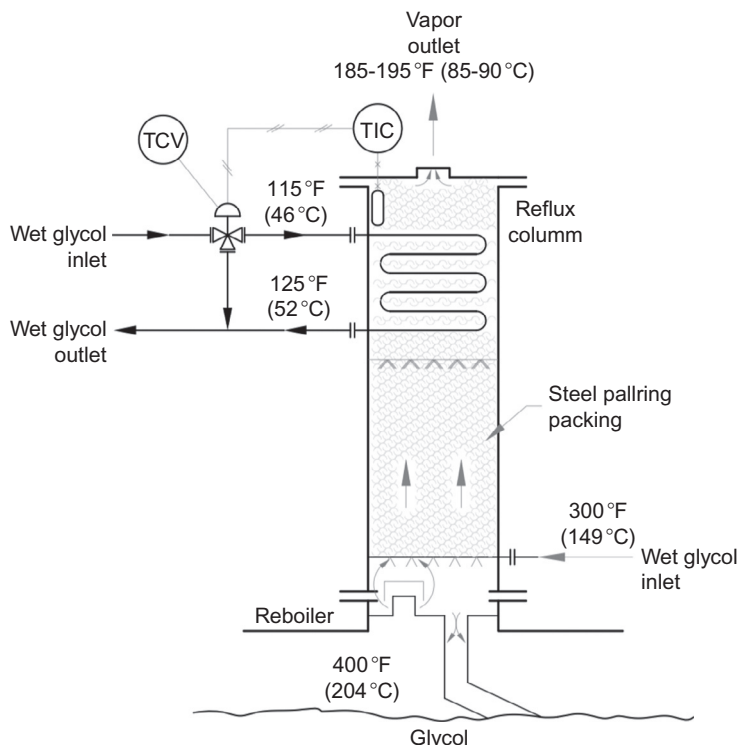


Figure 7.50 Still column with wet glycol entering below the stainless steel pall rings.

through the still column. The diameter required for the still is based on the glycol circulation rate (Figure 7.51).

7.5.18 Packing

One to three theoretical trays (4–12 ft.) is sufficient for most TEG stripping still requirements. 304 SS packing is normally used.

7.5.19 Amount of Stripping Gas

The amount of stripping gas required to reconcentrate the glycol to a high purity ranges from 2 to 10 ft.³/gal TEG circulated (Figure 7.52).

7.5.20 Filters

7.5.20.1 Microfiber

The microfiber filters are used to remove solid particles. They are typically sized to remove solids 5 μm and above.

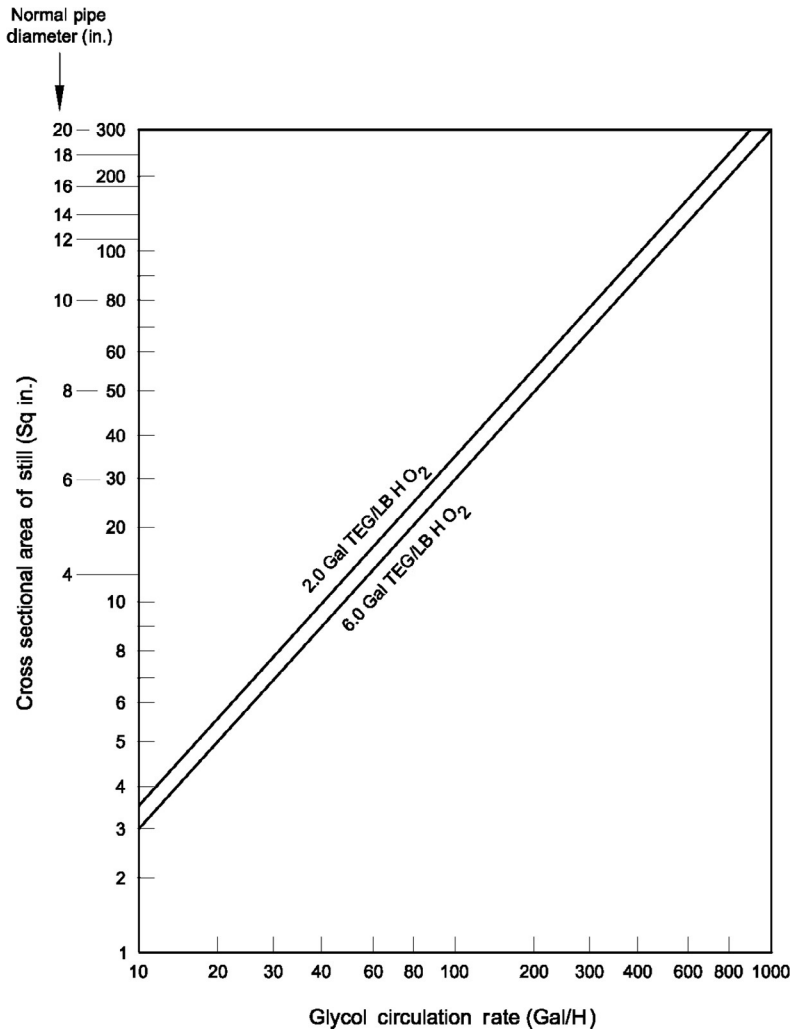


Figure 7.51 Determination of stripping still column diameter.

7.5.20.2 Activated Charcoal (Carbon)

Activated charcoal is used to remove chemical impurities. It is sized for full flow with 10 gpm streams and sized for 10-25% side streams on large units.

7.5.21 Glycol Pumps

Two types of pumps are used: Glycol-gas powered pumps and electricity-driven positive displacement piston/plunger pumps.

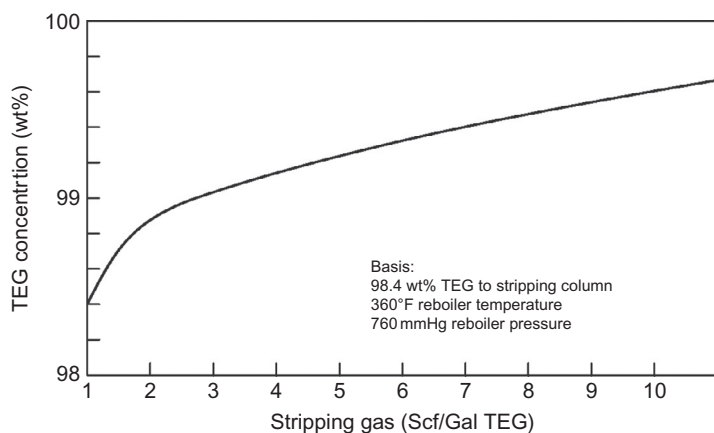


Figure 7.52 Amount of stripping gas required to reconcentrate glycol to high purity.

7.5.21.1 Glycol-Gas Powered Pumps (Figure 7.53)

Glycol-gas powered pumps are powered by gas entrained in the wet glycol leaving the contactor. They do not require contactor glycol liquid level control, dump valve, or external power (electricity). Gas consumption is relatively low and when used in conjunction with a glycol hydrocarbon skimmer or flash tank, very little gas loss is experienced. They have few moving parts, which translates to less wear and more simplified repairs. Contact with hydrocarbon distillate, which may be entrained in glycol passing through the pump, swells o-ring seals in the pumps causing premature pump failure. Glycol-gas powered pumps are typically used on small isolated systems. They are inexpensive and break often, but are easy to repair.

Circulation rate and gas consumption of the Kimray glycol-gas powered pump are shown in [Tables 7.5](#) and [7.6](#). Temperatures above 200 °F damage o-ring seals.

7.5.21.2 Electricity-Driven Positive Displacement Piston/Plunger Pump

Electricity-driven piston/plunger pumps are typically used in large installations. They require a small glycol leak in the piston rod packing for lubrication. They also are resilient to hydrocarbon distillate, grit, and debris that would damage the glycol-gas powered pumps.

7.5.22 Still Emissions

Vapor from the still column can contain some hydrocarbon gases that flashed from the glycol, stripping gas, and aromatics. Glycol preferentially absorbs aromatics and naphthene components over paraffinic components in the inlet gas.

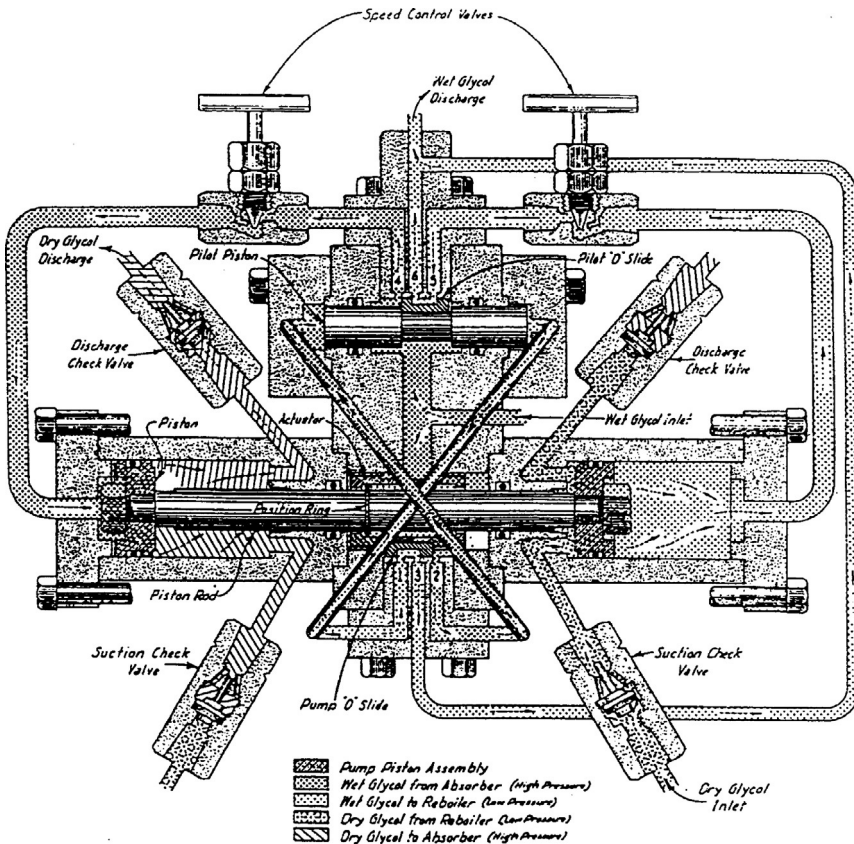


Figure 7.53 Operation of the glycol-gas powered pump (Kimray).

Aromatics include benzene, ethylene, toluene, and xylene (commonly called BETX). Aromatics condense with water vapor or can lead to “soluble” oil in the produced water discharge. Treatment consists of condensing the water vapor and BETX exiting the still column and then compressing the noncondensable (hydrocarbon gases) (Figure 7.54).



7.6 MERCURY CONSIDERATIONS

7.6.1 Mercury

Mercury can be present as elementary mercury or it can be introduced into the gas stream from instruments. Mercury has an affinity for higher molecular weight components. Most mercury stays with liquids rather than

Table 7.5 Glycol-Gas Circulation Rate (gallons/hour)

Model Number	Pump Speed (strokes/min) (Count One Stroke for Each Discharge of Pump)																
	8	10	12	14	16	18	20	22	24	26	28	30	32	34	36	38	40
1715V		10	12	14	16	18	20	22	24	26	28	30	32	34	36	38	40
4015V			12	14	16	18	20	22	24	26	28	30	32	34	36	38	40
9015V	8		27	31.5	36	40.5	45	49.5	54	58.5	63	67.5	72	76.5	81	85.5	90
21015V		66	79	92	105	118	331	144	157	171	184	19	210				
45015V		166	200	233	266	300	333	366	400	433	197						

Note: Do not attempt to run pumps at speeds less or greater than those indicated above.

Table 7.6 Gas Consumption

Operating Pressure (psig)	300	400	500	600	700	800	900	1000	1100	1200	1300	1400	1500
Gas Consumption (SCF/Gallon)	1.7	2.3	2.8	3.4	3.9	4.5	5.0	5.6	6.1	6.7	7.2	7.9	8.3

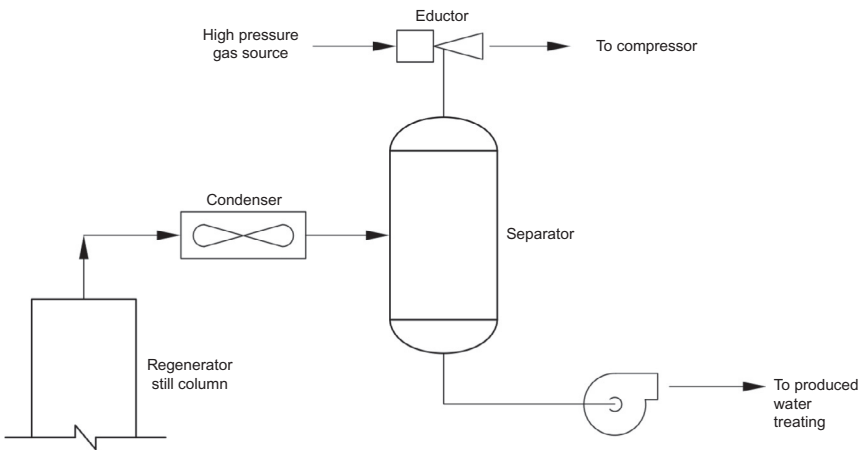
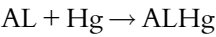


Figure 7.54 Process flow diagram of treatment of still emissions.

with the gas stream. Mercury reacts with iron oxide in the presence of hydrogen (H₂) and deposits mercurous sulfide on carbon steel pipe walls. In the presence of condensed water, mercury combines with aluminum to form a weak amalgam. Mercury has a cumulative effect, thus even trace amounts can be harmful.



and



7.6.2 Treatment

Mercury is treated with a sulfur-impregnated activated carbon with at least a 10% sulfur content. System design considerations are as follows:

Bed absorption = 15–20 wt% Hg

Pressure range = 300–1100 psi

Temperatures = up to 175 °F

Gas contact time = 20 s

Maximum velocity = 35 fpm

Regeneration = no commercial process

Rhone-Puolene offers proprietary treatment with alumina beds that can be recycled.



7.7 SPECIAL GLYCOL DEHYDRATION SYSTEMS

7.7.1 General Considerations

When large dew-point depressions are required, specialized dehydration systems using highly concentrated glycol is necessary. If space is limited, special systems may be used to achieve the required dew-point depressions. Normal dehydration systems with TEG glycol purity of 98.5% are capable of achieving dew-point depressions up to 70 °F. Stripping gas may be used to obtain higher dew-point depressions. Vacuum-operated glycol units can achieve glycol purities of up to 99.9%, but they are rarely used because of high operating costs and problems associated with achieving the necessary vacuums.

The Drizo process and the Cold-Finger Condenser process are other methods used to obtain low dew points.

7.7.2 Drizo Process

The Drizo process is used to obtain glycol concentrations (purity) as high as 99.99% and low dew points in the range of –40 to –80 °F. This process uses an 80 to 100 molecular weight solvent (usually 150 octane) in the reconcentrator to form an azeotrope with water, thus lowering the effective boiling point of the mixture. Glycol of higher purity is thus achieved for a given reconcentrator temperature. BETX are collected as “excess” solvents. The Drizo process may be favored over stripping gas, and existing units can be retrofitted to increase the dehydration capacity. Each situation must be evaluated on a case-by-case basis because the Drizo process is a Dow-patented process requiring a license fee for use.

7.7.2.1 Process Description (*Figure 7.55*)

As shown in [Figure 7.55](#), the Drizo process is the same as a conventional TEG dehydration system until the wet glycol flows into the reconcentrator.

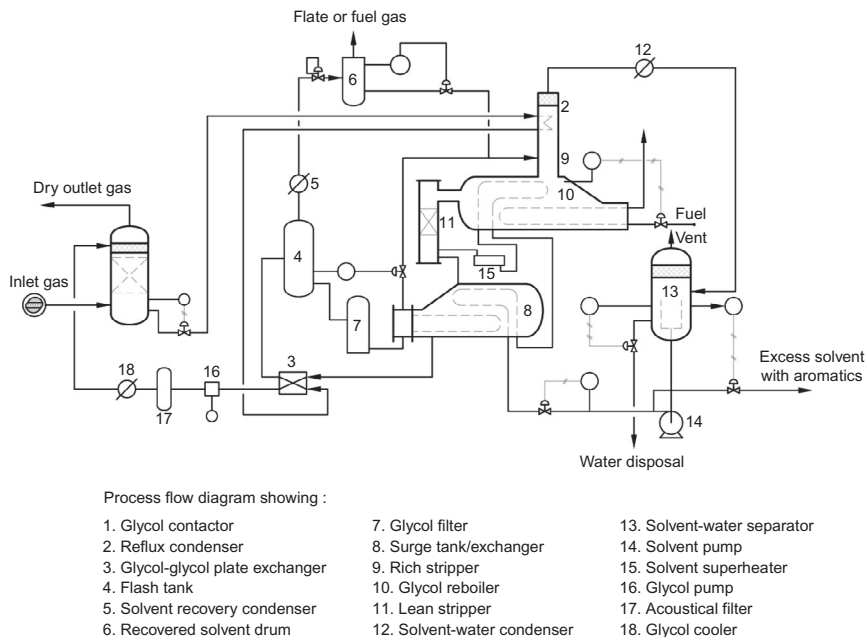


Figure 7.55 Dow Drizo gas dehydration process.

Wet glycol is reconcentrated to 98.5% by conventional distillation. The semi-lean glycol is then counter-currently contacted with hydrocarbon solvent (iso-octane) vapors at 400 °F. The hydrocarbon and water are taken overhead, condensed, and then phase separated. The water is discarded, and the solvent is recycled into the system.

7.7.2.2 Applications

The Drizo process is competitive with applications that utilize a conventional TEG unit with stripping gas. It is most competitive in the range of -40 to -80 °F. It should not be considered for hydrocarbon dew points.

7.7.3 Cold-Finger Condenser Process

The Cold-Finger Condenser process is based on the water TEG, vapor liquid equilibrium diagram shown in [Figure 7.56](#). This diagram shows that for any liquid concentration, the corresponding equilibrium vapor concentration is richer in water. The process incorporates a closed vessel half-filled with vapor and liquid at equilibrium with a condenser-tube bundle in the vapor space ([Figure 7.57](#)). The condenser causes water condensation, which

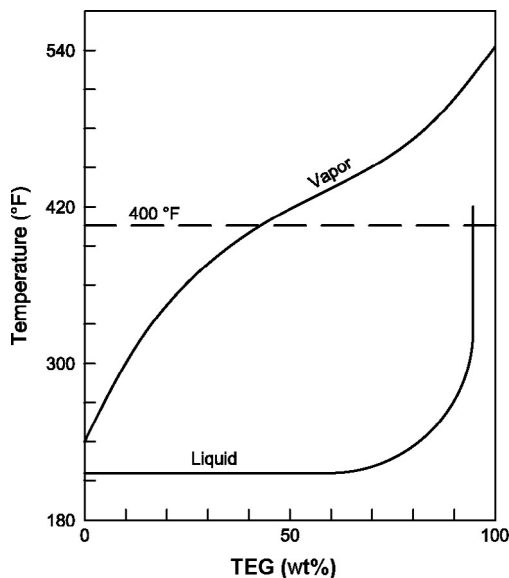


Figure 7.56 Vapor liquid equilibrium diagram for TEG.

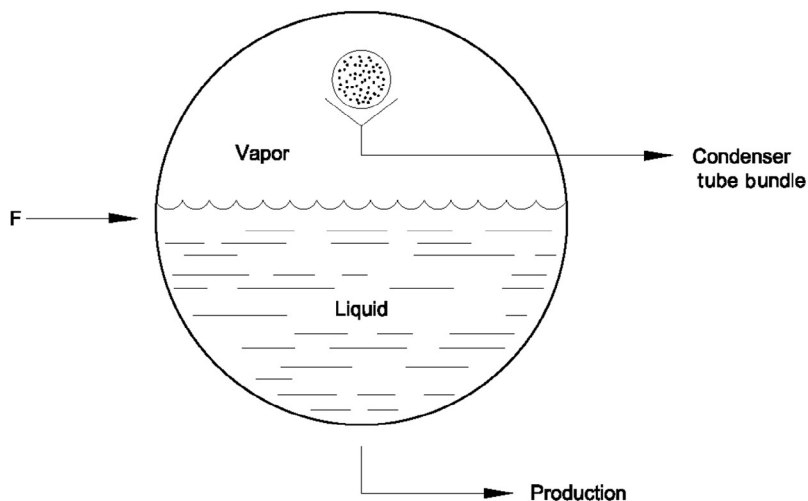


Figure 7.57 Cold-finger condenser.

is removed from the vessel to a trough placed under the condenser-tube bundle. As the condensate is removed, the system's equilibrium is upset and the liquid phase releases more water to the vapor in order to reestablish equilibrium. Consequently, the liquid phase has a lower water content than it did originally.

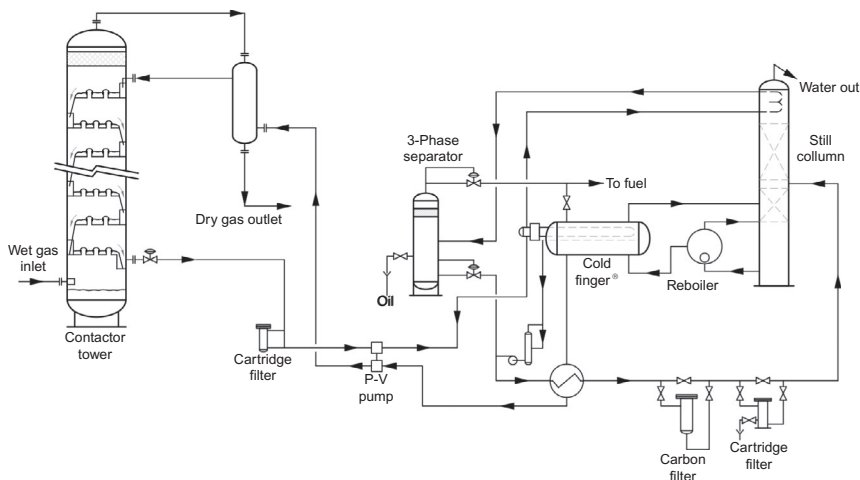


Figure 7.58 Cold-finger condenser process.

7.7.3.1 Process Description

There are numerous variations based on this principle. One design is shown in [Figure 7.58](#). Contact between gas and glycol is the same as in a conventional TEG system. Wet glycol leaves the contractor and flows to the condenser-tube bundle of the cold finger, where it acts as a coolant and is used as a coolant in the glycol still before the hydrocarbon liquid phase, hydrocarbon vapor phase, and glycol/water phase are separated in a three-phase separator. The glycol/water phase is mixed with the cold-finger condensate, and is heated by the cold-finger liquid product before it is fed to the still. The hot semi-lean glycol (which is near its boiling point) from the still bottoms is fed to the cold finger. The liquid product is cooled, pumped, cooled again, and fed to the contactor.

7.7.3.2 Application

The main benefit of this system is that it is more fuel-efficient than the conventional TEG system. However, it is also more complex and not as well-proven as the conventional system.

7.8 SYSTEMS UTILIZING GLYCOL-GAS POWERED PUMPS ([FIGURE 7.59](#))

Cool wet glycol leaves the bottom of the contactor, passes through a strainer, and powers the pump. The wet glycol takes a pressure drop through

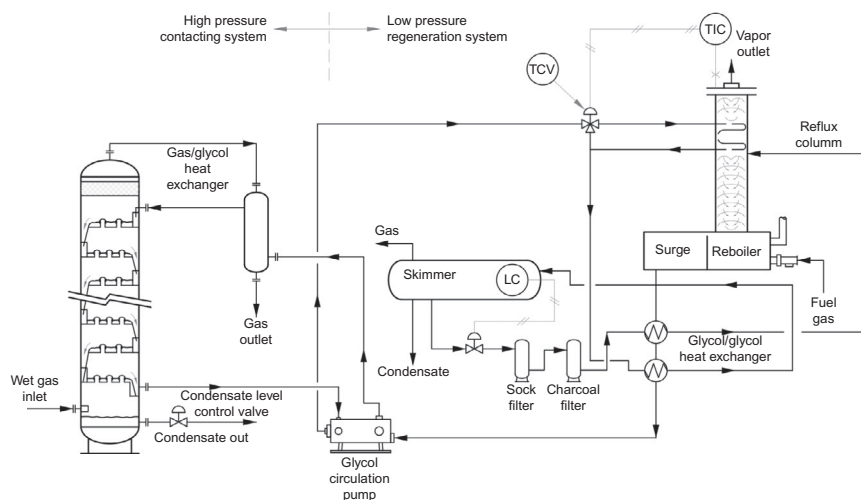


Figure 7.59 System utilizing glycol-gas powered pumps.

the pump, then passes through the reflux condenser coil in the reconcentrator still column.



7.9 SYSTEMS UTILIZING ELECTRICITY-DRIVEN PUMPS

Cool wet glycol leaves the bottom of the contactor, passes through a choke and level-control operated motor valve, where a pressure drop occurs. The glycol then passes through the still column reflux coils. From the reflux coil, it flows through the first dry glycol to wet glycol heat exchanger, then into the gas-glycol-condensate separator, where insoluble hydrocarbons are removed. From the separator, the glycol passes through the filter to remove tarry hydrocarbons. It then goes through the second dry glycol to wet glycol heat exchanger and into the reconcentrator still column. In the top of the reboiler still column, cool wet glycol flows through the reflux condenser coils, preventing glycol from leaving as a vapor (see [Figure 7.60](#)).

The wet glycol enters the column below the coils and spills downward through the packing and into the reconcentrator. Heat is circulated through the tube to boil the water from the glycol. A weir holds the level of glycol above the heating tubes. Regenerated glycol flows over the weir and leaves through the outlet in the bottom.

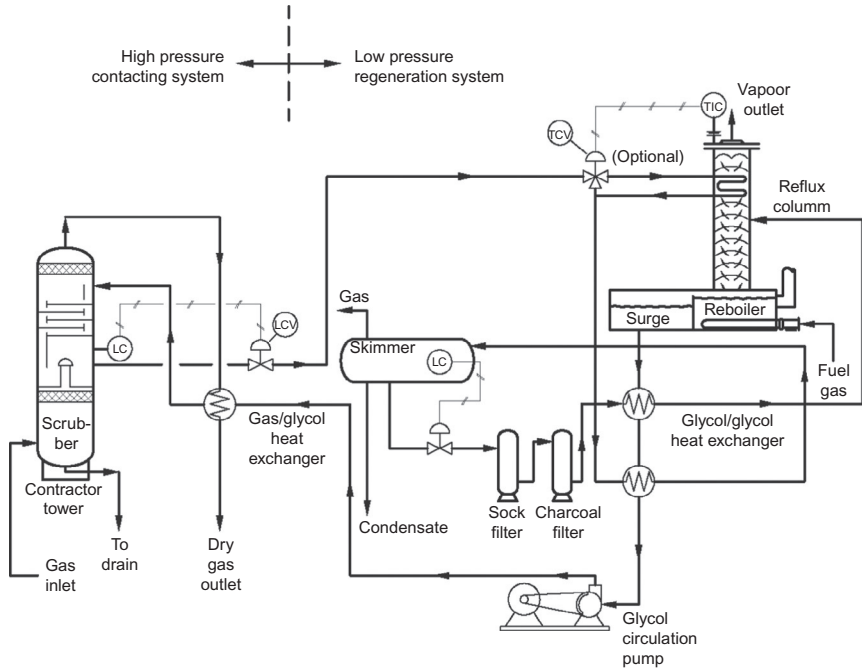


Figure 7.60 Systems utilizing electricity-driven pumps.

Example 7.3 Glycol Dehydration

Given

Gas $Q_g = 98$ MMSCFD at 0.67 SG saturated with water at 1000 psig and 100 °F

Dehydrate to = 7 lb/MMSCF

Use TEG

No stripping gas is available

98.5% TEG concentration

C_D (contactor) = 0.852

$T_C = 376$ °R

$P_C = 669$ psia

Determine

1. Calculate contactor diameter
2. Determine glycol circulation rate and estimate reboiler duty
3. Size the still column
4. Calculate heat duties for gas/glycol exchanger and glycol/glycol exchangers

Solution

1. Calculate contactor diameter

$$d^2 = 5040 \left(\frac{TZQ_g}{P} \left[\left(\frac{\rho_g}{\rho_L \rho_t} \right) \frac{C_D}{d_M} \right] \right)^{1/2}$$

$$\begin{aligned} d_M &= 125 \text{ } \mu\text{m} \quad (\text{range } 120\text{--}150 \text{ } \mu\text{m}); \quad T = 570 \text{ } ^\circ\text{R}; \quad P = 1015 \text{ psia}; \\ Q_g &= 98 \text{ MMSCFD}; \quad T_r = 570/376 = 1.49; \quad P_r = 1015/669 = 1.52; \\ Z &= 0.865 \end{aligned}$$

$$\begin{aligned} \rho_g &= \frac{[(0.67)(1015)]}{[(560)(0.865)]} \\ &= 3.79 \text{ lb/ft.}^3 \end{aligned}$$

$$\rho_L = 70 \text{ lb/ft.}^3; \quad C_D = 0.852 \text{ (given)}$$

$$\begin{aligned} d^2 &= 5040 \frac{(560)(0.685)(98)}{(1015)} \left[\left(\frac{3.72}{70-3.79} \right) \frac{0.852}{125} \right]^{1/2} \\ &= 68.2 \text{ in.} \end{aligned}$$

Use 72" ID contactor (standard off-the-shelf)

2. Determine glycol circulation rate and reboiler duty

$$\begin{aligned} W_i &= 63 \text{ lb/MMSCF (from McKetta-Wehe) (saturated water content);} \\ W_0 &= 7 \text{ lb/MMSCF (spec); } \Delta W = W_i - W_0 = 63 - 7 = 56 \text{ lb/MMSCF;} \\ \Delta W/W_i &= 56/63 = 0.889. \end{aligned}$$

Using $n=2$ (i.e., eight actual trays) and glycol purity of 98.5%, read from [Figure 7.48](#), the glycol circulation rate of 2.8 gal TEG/lb H₂O. Use 3.0 gal/lb for design.

$$\begin{aligned} L &= \left(\frac{2.0 \text{ gal}}{\text{lb}} \right) \left(\frac{56 \text{ lb}}{\text{MMSCF}} \right) \left(\frac{98 \text{ MMSCF}}{D} \right) \left(\frac{D}{24 \text{ h}} \right) \left(\frac{\text{h}}{60 \text{ min}} \right) \\ &= 11.4 \text{ gpm TEG} \\ &= 862 \text{ Btu/gal (table)} \\ &= \left(\frac{862 \text{ Btu}}{\text{gal}} \right) \left(\frac{11.4 \text{ gal}}{\text{min}} \right) \left(\frac{60 \text{ min}}{\text{h}} \right) \\ &= 590 \text{ Mbtu/h} \end{aligned}$$

To allow for start-up heat loads, increase heat duty by 10% and then select a standard off-the-shelf fire tube. Thus, select a 750 MMBtu/h.

3. Design of still column

Use a 12-foot still column (standard packed arrangement)

$$d_M = 125 \text{ } \mu\text{m}; \quad T = 300 \text{ } ^\circ\text{F} = 760 \text{ } ^\circ\text{R}; \quad P = 1 \text{ psig}$$

$$Q_g = \left(\frac{10 \text{ scf}}{\text{gal}} \right) \left(\frac{11 \text{ gal}}{\text{min}} \right) \left(\frac{60 \text{ min}}{\text{h}} \right) \left(\frac{24 \text{ h}}{d} \right)$$

$$= 0.16 \text{ MMSCFD}$$

$$Z = 1.0$$

$$\rho_g = 2.7 \left(\frac{(0.62)(16)}{(760)(1.0)} \right)$$

$$= 0.035 \text{ lb/ft}^3$$

$$\rho_L = 62.4 \text{ lb/ft}^3; C_D = 14.2 \text{ (given)}$$

$$d^2 = 5040 \frac{(760)(1.0)(0.16)}{16} \left[\left(\frac{0.035}{62.4 - 0.035} \right) \frac{14.2}{125} \right]^{1/2}$$

$$= 17.5 \text{ in.}$$

Use 18 in. OD \times 12 ft. long still

4. Calculate duties of heat exchangers

Rich TEG from contactor: $T = 100^\circ\text{F}$ (given)

Rich TEG to separator: $T = 200^\circ\text{F}$ (assume for good design)

Rich TEG from reflux: $T = 110^\circ\text{F}$ (assume 10°F increase in reflux coil)

Rich TEG to still: $T = 300^\circ\text{F}$ (assume for good design)

Lean TEG from reboiler: $T = 385^\circ\text{F}$ (from [Figure 7.59](#))

Lean TEG to pumps (max): $T = 210^\circ\text{F}$ (from manufacturer)

Lean TEG to contactor: $T = 110^\circ\text{F}$ (10°F above contactor temperature)

5. Glycol/glycol preheater (rich side, duty):

Rich TEG: $T_1 = 110^\circ\text{F}$ (assume 10°F increase in reflux coil)

$T_2 = 200^\circ\text{F}$

Lean glycol composition:

$$W_{\text{TEG}} = (0.985) \left(\frac{70 \text{ lb}}{\text{ft}^3} \right) \left(\frac{\text{ft}^3}{7.48 \text{ gal}} \right) W_{\text{TEG}}$$

$$= 9.22 \text{ lb TEG/gal of lean glycol}$$

$$W_{\text{H}_2\text{O}} = (0.015) \left(\frac{70 \text{ lb}}{\text{ft}^3} \right) \left(\frac{\text{ft}^3}{7.48 \text{ gal}} \right)$$

$$= 0.140 \text{ lb H}_2\text{O/gal of lean glycol}$$

Rich glycol composition

$W_{\text{TEG}} = 9.22 \text{ lb TEG/gl of lean glycol}$

$$W_{H_2O} = \left(\frac{0.140 \text{ lb } H_2O}{\text{gal of lean glycol}} \right) + \left(\frac{1 \text{ lb } H_2O}{3.0 \text{ gal of lean glycol}} \right)$$

$$= 0.473 \text{ lb } H_2O / \text{gal of lean glycol}$$

$$\text{Wt. Concentration TEG} = \frac{9.22}{9.22 + 0.473}$$

$$= 95.1\%$$

Rich glycol flow rate (W_{rich})

$$W_{\text{rich}} = (9.22 + 0.473) \frac{\text{lb}}{\text{gal}} \left(\frac{11.4 \text{ gal}}{\text{min}} \right) \left(\frac{60 \text{ min}}{\text{h}} \right)$$

$$= 6630 \text{ lb/h}$$

Rich glycol heat duty (q_{rich})

C_p (95.1% TEG) = 0.56 at 110 °F (from physical property of TEG)
and 0.63 at 200 °F = 0.60 Btu/h °F $C_{p, \text{AVG}}$ = 0.60 Btu/h °F

$$q_{\text{rich}} = \left(\frac{6630 \text{ lb}}{\text{h}} \right) \left(\frac{0.6 \text{ btu}}{\text{h}} \right) (200 - 110)^\circ\text{F}$$

$$= 358 \text{ MBtu/h}$$

6. Glycol/glycol exchanger

Rich $T_1 = 200$

$T_2 = 300$

Lean $T_3 = 390$

$T_4 = ?$

Rich glycol heat duty:

C_p (95.1% TEG) = 0.63 at 200 °F (from physical properties of TEG) =
0.70 at 300 °F $C_{p, \text{AVG}}$ = 0.67 Btu/h °F

$$q_{\text{rich}} = \left(\frac{6630 \text{ lb}}{\text{h}} \right) \left(\frac{0.6 \text{ btu}}{\text{h}} \right) (300 - 200)^\circ\text{F}$$

$$= 444 \text{ MBtu/h}$$

$$W_{\text{lean}} = \left(\frac{11.4 \text{ gal}}{\text{min}} \right) \left(\frac{70 \text{ lb}}{\text{ft}^3} \right) \left(\frac{\text{ft}^3}{7.84 \text{ gal}} \right) \left(\frac{60 \text{ min}}{\text{h}} \right)$$

$$= 6401 \text{ lb/h}$$

Calculation of T_4

Assume $T = 250^\circ\text{F}$

$$T_{\text{AVG}} = (353 + 250)/2 = 302^\circ\text{F}$$

$C_{p,\text{AVG}} = (98.5\% \text{ TEG}) = 0.67 \text{ Btu/lb }^\circ\text{F}$ (from physical properties of TEG)

$$Q_{\text{lean}} = W_{\text{lean}} C_p (T_4 - T_3)$$

$$Q_{\text{lean}} = -q_{\text{rich}}$$

$$\begin{aligned} T_4 &= T_3 - \left(\frac{q_{\text{rich}}}{W_{\text{lean}} C_p} \right) \\ &= 353 - \left(\frac{444,000}{(6401)(0.67)} \right) \\ &= 249^\circ\text{F} \end{aligned}$$

Temperature:

Lean: $T_4 = 249^\circ\text{F}$

$T_5 = ?$

Assume $T_5 = 175^\circ\text{F}$

$$T_{\text{AV}} = (249 + 175)/2 = 212$$

$C_{p,\text{AV}} = (98.5\% \text{ TEG}) = 0.61 \text{ Btu/lb }^\circ\text{F}$ (from physical properties of TEG)

$$q_{\text{lean}} = W_{\text{lean}} C_p (T_4 - T_5)$$

$$q_{\text{lean}} = -q_{\text{lean}}$$

$$\begin{aligned} T_5 &= T_4 - \left(\frac{q_{\text{rich}}}{W_{\text{lean}} C_p} \right) \\ &= 249 - \left(\frac{358,000}{(6401)(0.61)} \right) \end{aligned}$$

$= 157^\circ\text{F}$ (this is less than the maximum allowed to the pumps)

Lean: $T_1 = 157^\circ\text{F}$

$T_2 = 110^\circ\text{F}$

C_p (98.5% TEG) = 0.57 at 157°F (from physical properties of TEG) = 0.53 at 110°F

$$C_{p,\text{AVG}} = 0.55 \text{ Btu/lb }^\circ\text{F}$$

$$q_{\text{lean}} = (6401) (0.55) (110 - 157) = -165 \text{ MBtu/h}$$

Glycol/glycol exchanger:

Rich $T_1 = 200^\circ\text{F}$, $T_0 = 300^\circ\text{F}$

Lean $T_2 = 353^\circ\text{F}$, $T_0 = 249^\circ\text{F}$

Duty $q = 444 \text{ MBtu/h}$



7.10 NONREGENERABLE DEHYDRATOR

7.10.1 Overview

One other general category is covered specifically, the calcium chloride brine dehydrator.

7.10.2 Calcium Chloride Unit

The calcium chloride (CaCl_2) dehydrator is the most common (Figure 7.61).

A unit typically consists of three sections:

Inlet gas scrubber

Brine tray

Solid brine particles

Only moving parts control the liquid levels for hydrocarbon liquid and the brine-water mixture.

7.10.3 Principles of Operation

Solid desiccant is placed in the top of the unit.

Water-wet gas containing the solid CaCl_2 gives up part of its water to form liquid brine to drip down and fill the trays.

Inlet gas, coming up through the specially designed nozzles on the trays, contact the brine efficiently.

The wettest gas contacts the most dilute brine (about 1.2 specific gravity).

Approximately 2.5 lb H_2O /lb CaCl_2 is removed in the trays.

Brine gravity on the top tray is about 1.4.

Another 1 lb H_2O /lb CaCl_2 is removed in the solid bed section.

Maximum dew-point depression of 60–70 °F occurs in this section.

Typically used in remote, small gas fields without heat or fuel.

Advantages

Simple

No moving parts

No heat required

Does not react with H_2S or CO_2

Can dehydrate hydrocarbon liquids

Disadvantages

Batch process

Emulsifies with oil

Unreliable

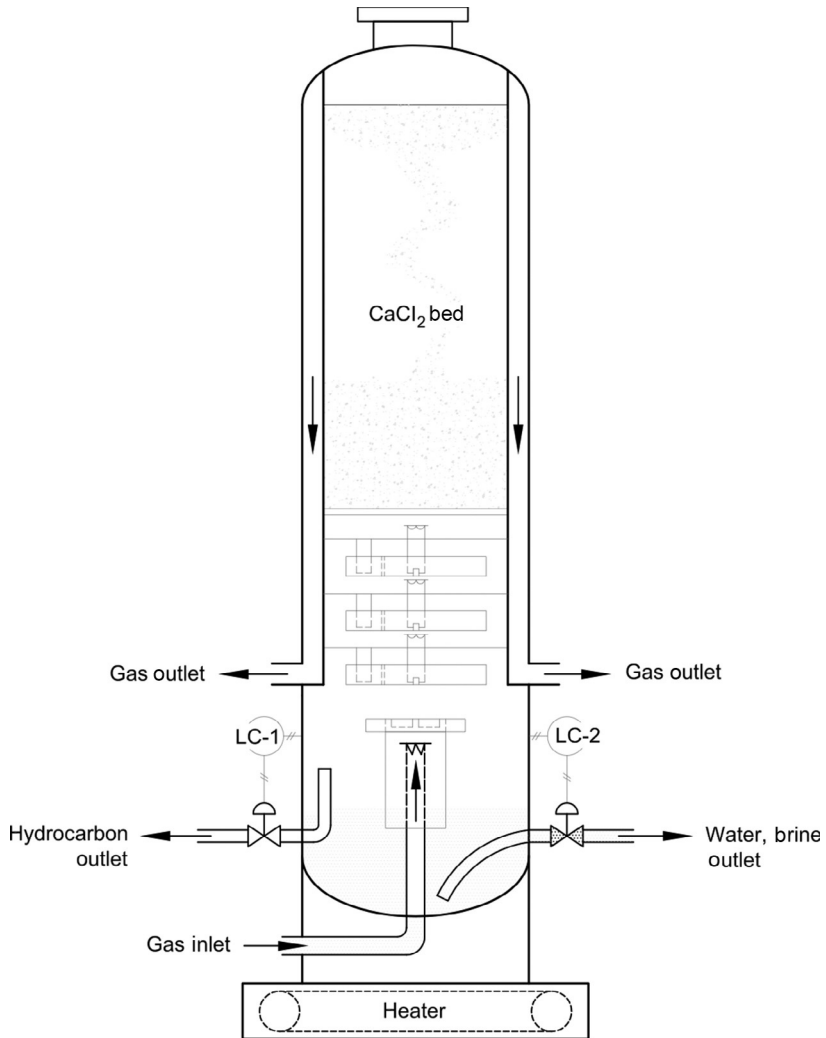


Figure 7.61 Cross section of calcium chloride dehydrator.

Limited dew-point depression
Brine disposal is a problem
Sensitive to varying flow rates

7.10.4 Operating Problems

Bridging and channeling is a problem.

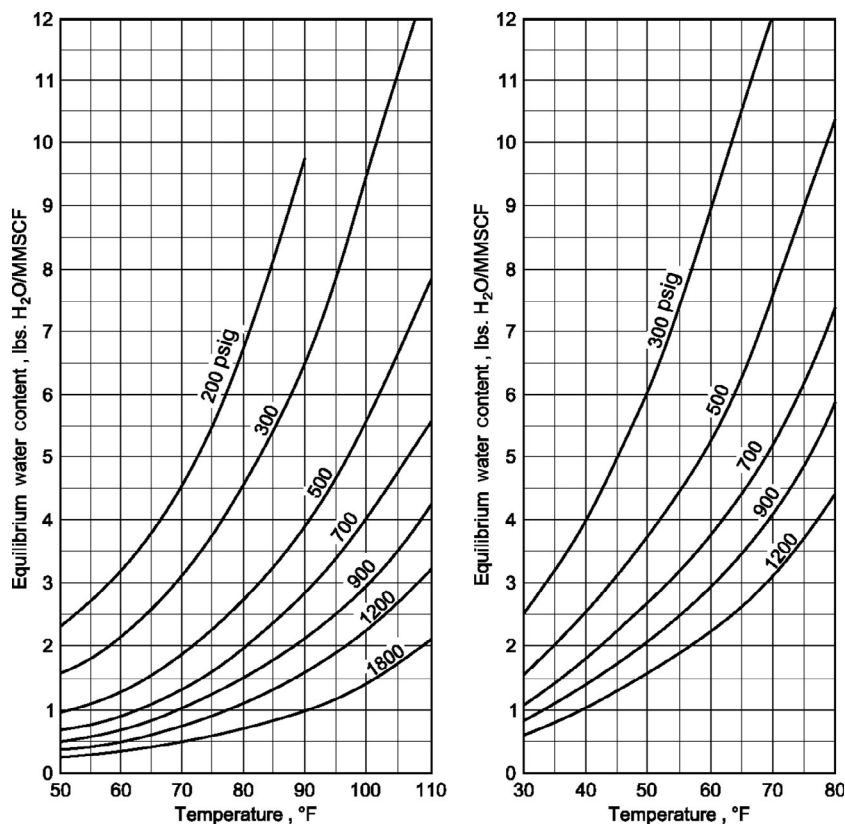


Figure 7.62 Approximate water content on natural gas driven by CaCl_2 unit (left: freshly recharged, right: just prior to recharging).

Brine can crystallize at 85 °F; thus, during low flow periods it can plug the vessel outlet or trays.

Brine carryover can cause severe corrosion problems.

7.10.5 Design Considerations

Figure 7.62 illustrates the water content of natural gas dried by solid calcium chloride bed units.



7.11 PHYSICAL PROPERTIES OF COMMON GLYCOLS

Figures 7.63–7.73 contain specific heats, specific gravities, and viscosities of EG, DEG, TEG and TTEG solutions.

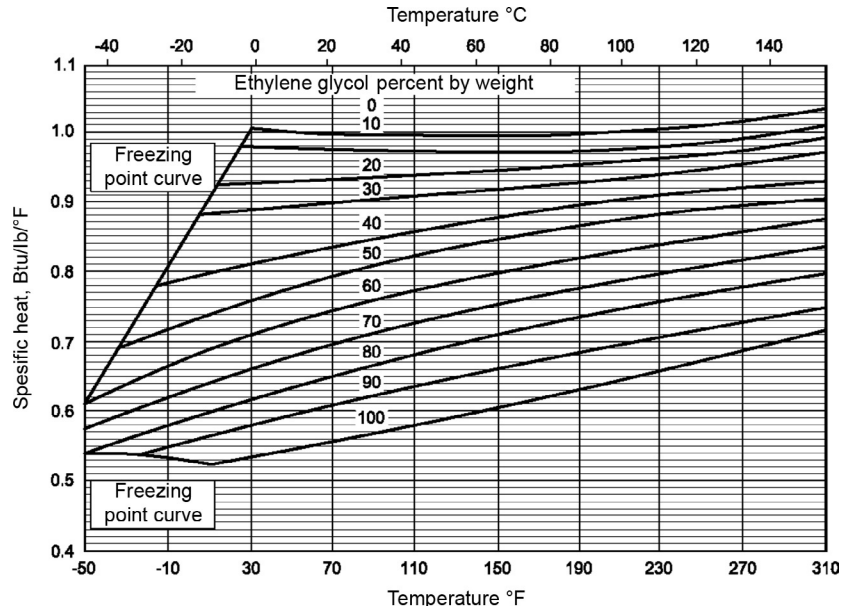


Figure 7.63 Specific heat of aqueous EG solutions.

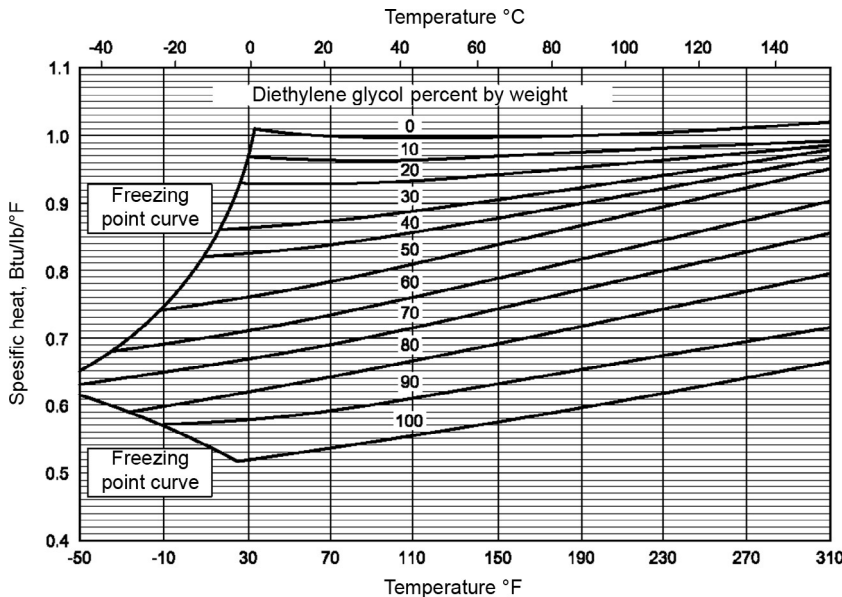


Figure 7.64 Specific heat of aqueous DEG solutions.

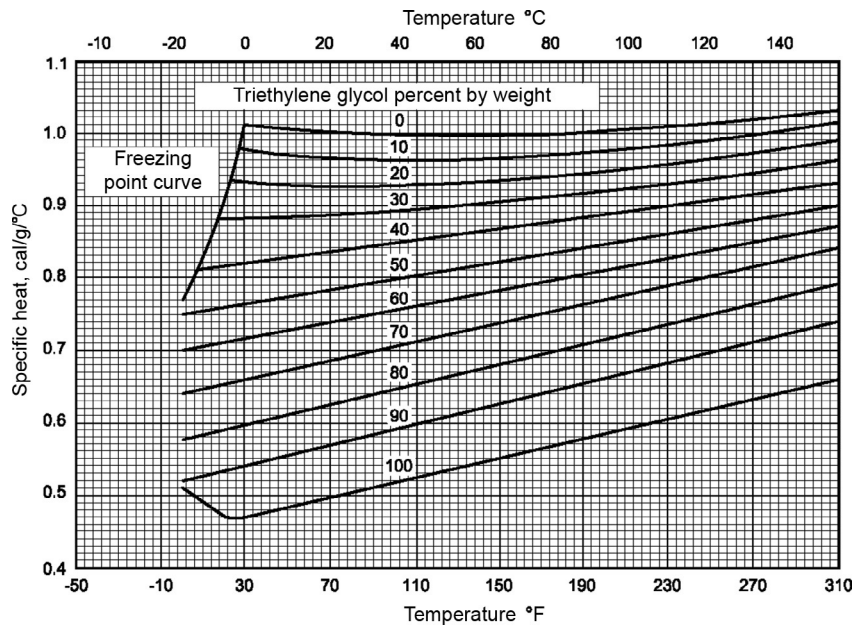


Figure 7.65 Specific heat of aqueous TEG solutions.

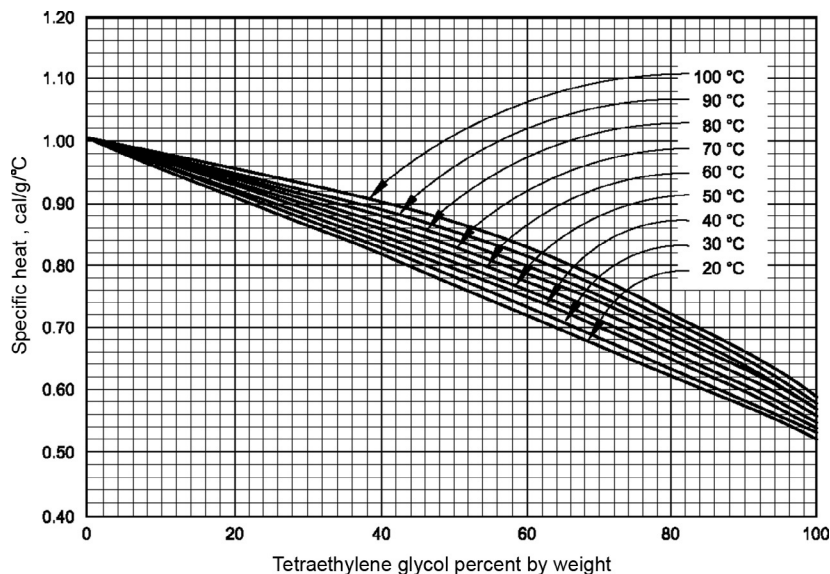


Figure 7.66 Specific heat of aqueous TTEG solutions.

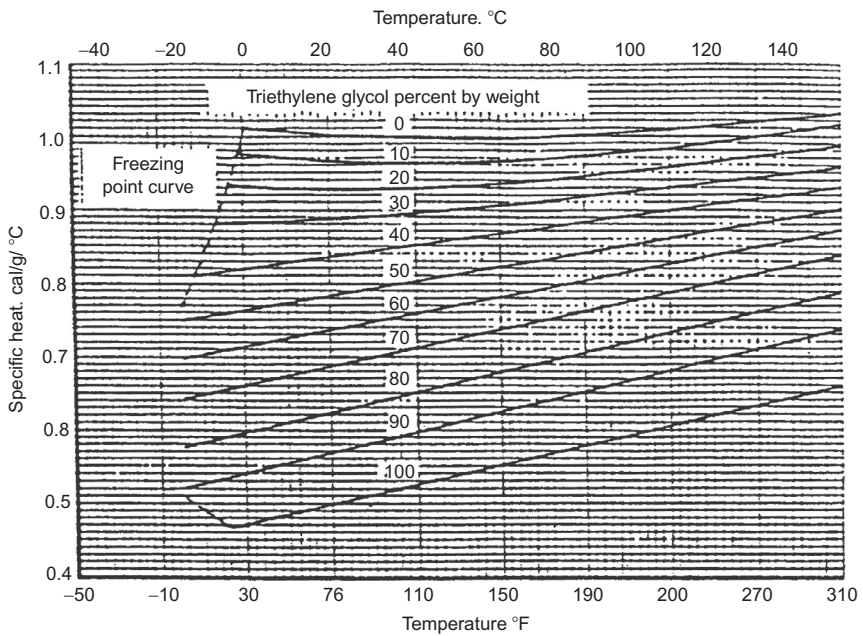


Figure 7.67 Specific gravity of aqueous TEG solutions.

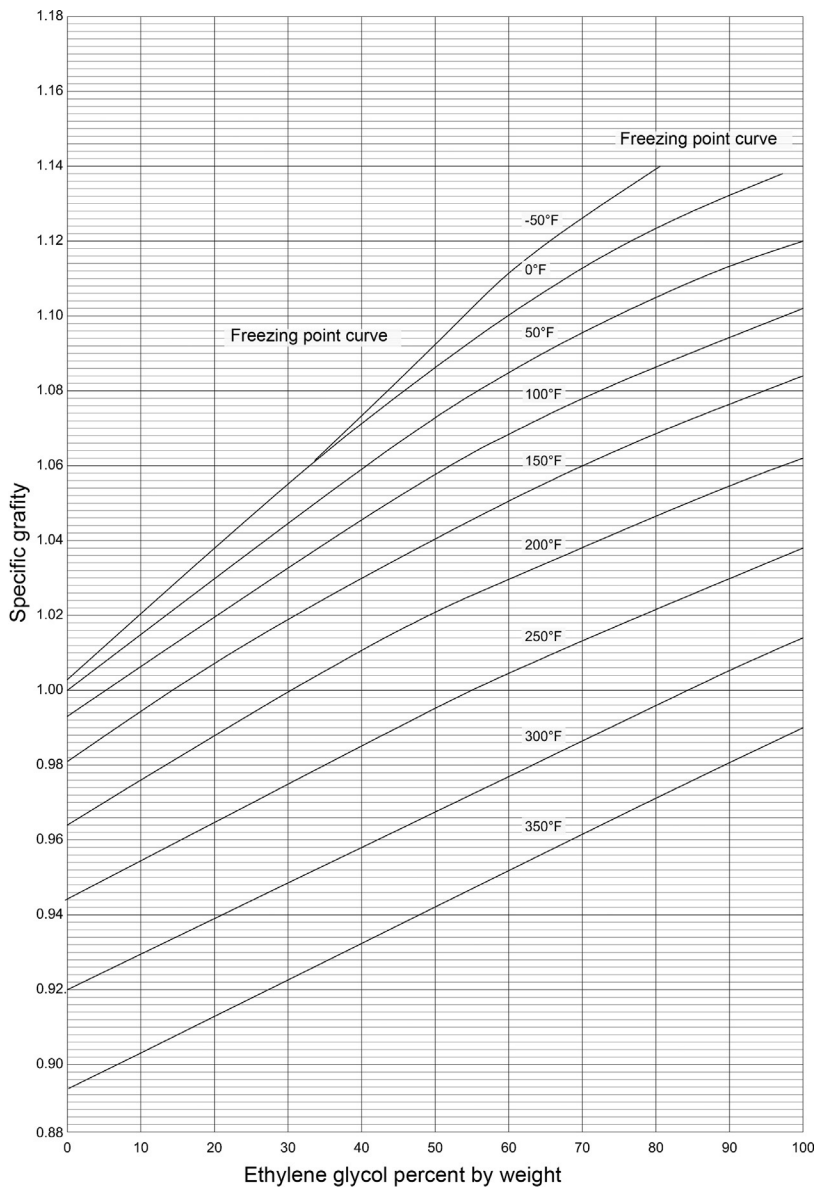


Figure 7.68 Specific gravity of aqueous EG solutions.

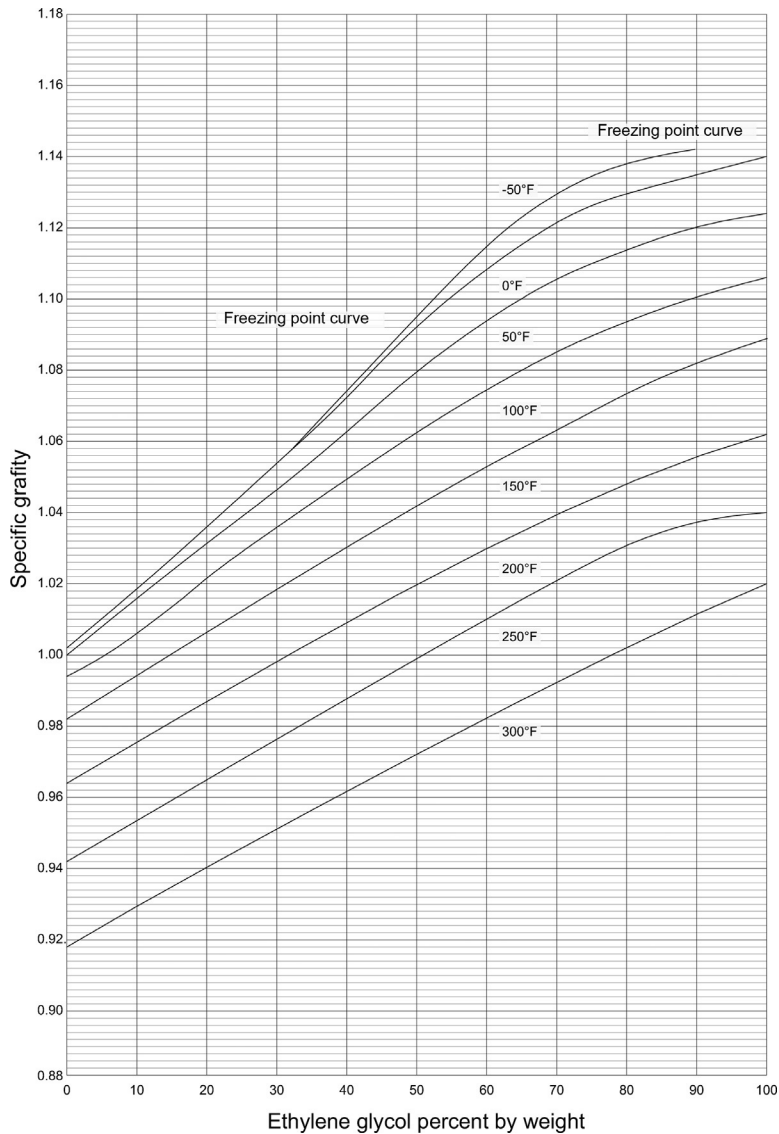


Figure 7.69 Specific gravity of aqueous DEG solutions.

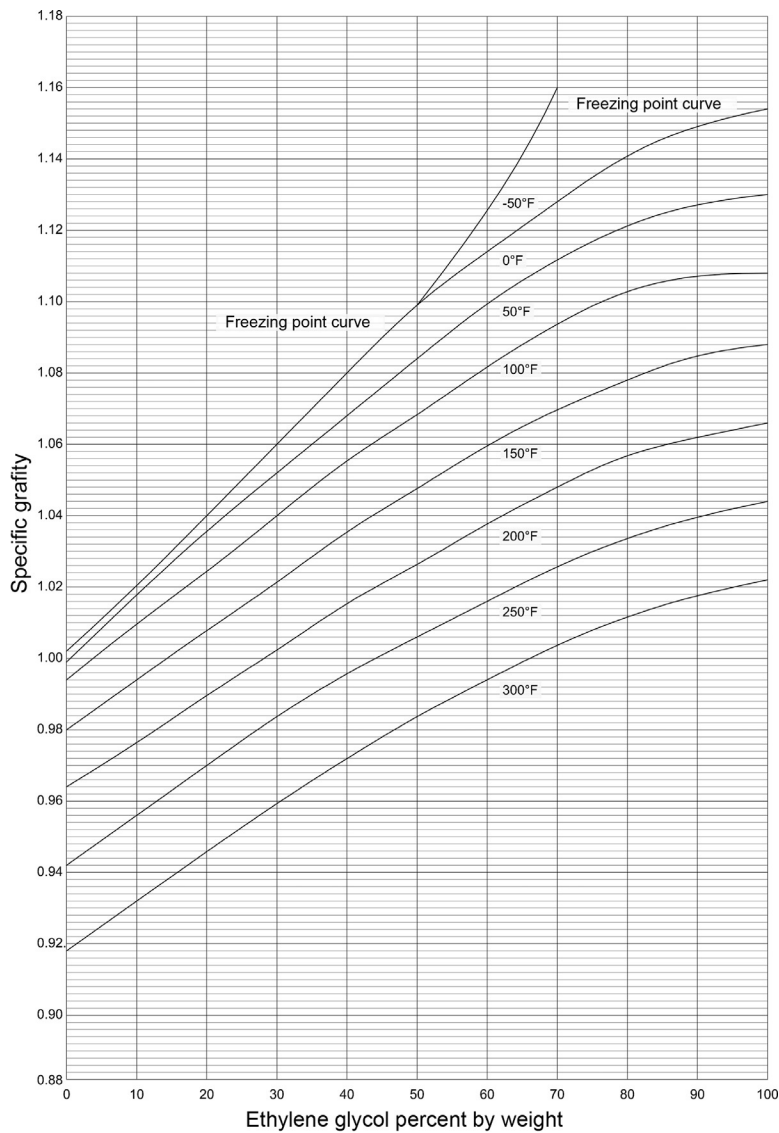


Figure 7.70 Specific gravity of aqueous TEG solutions.

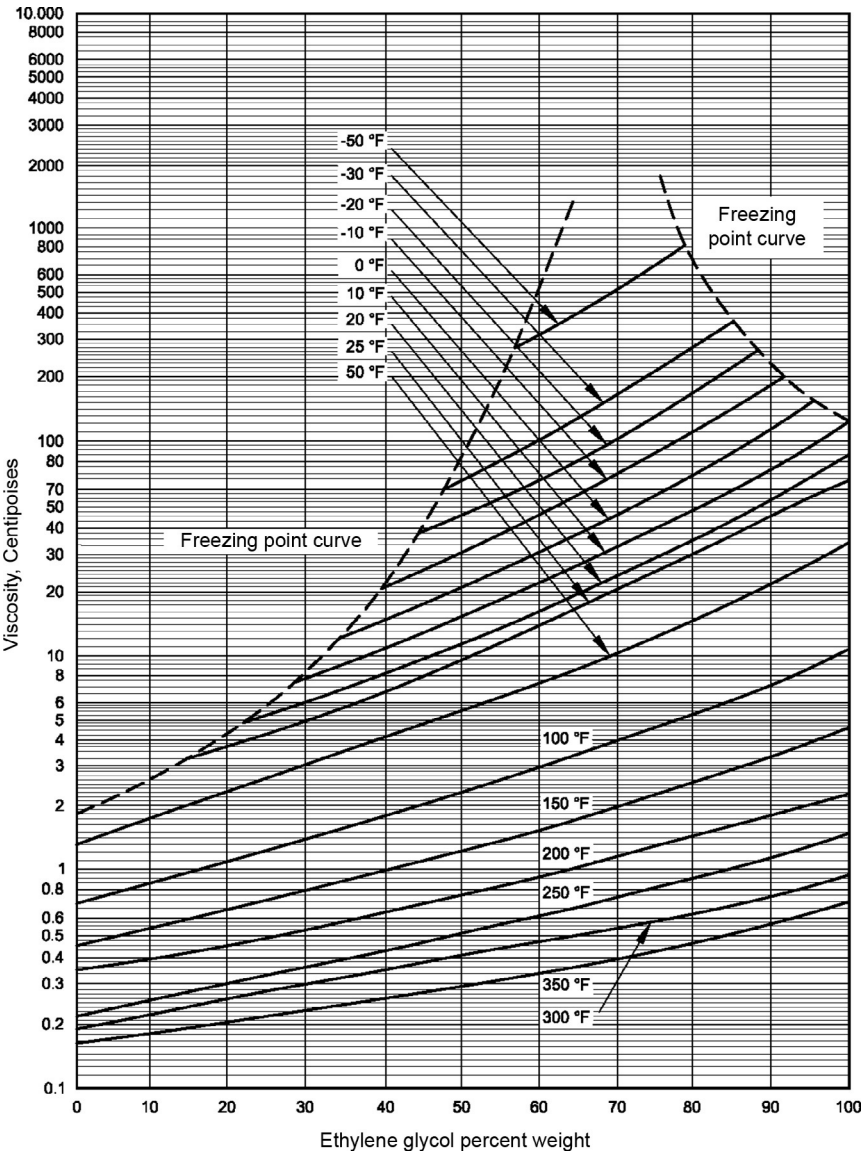


Figure 7.71 Viscosities of aqueous EG solutions.

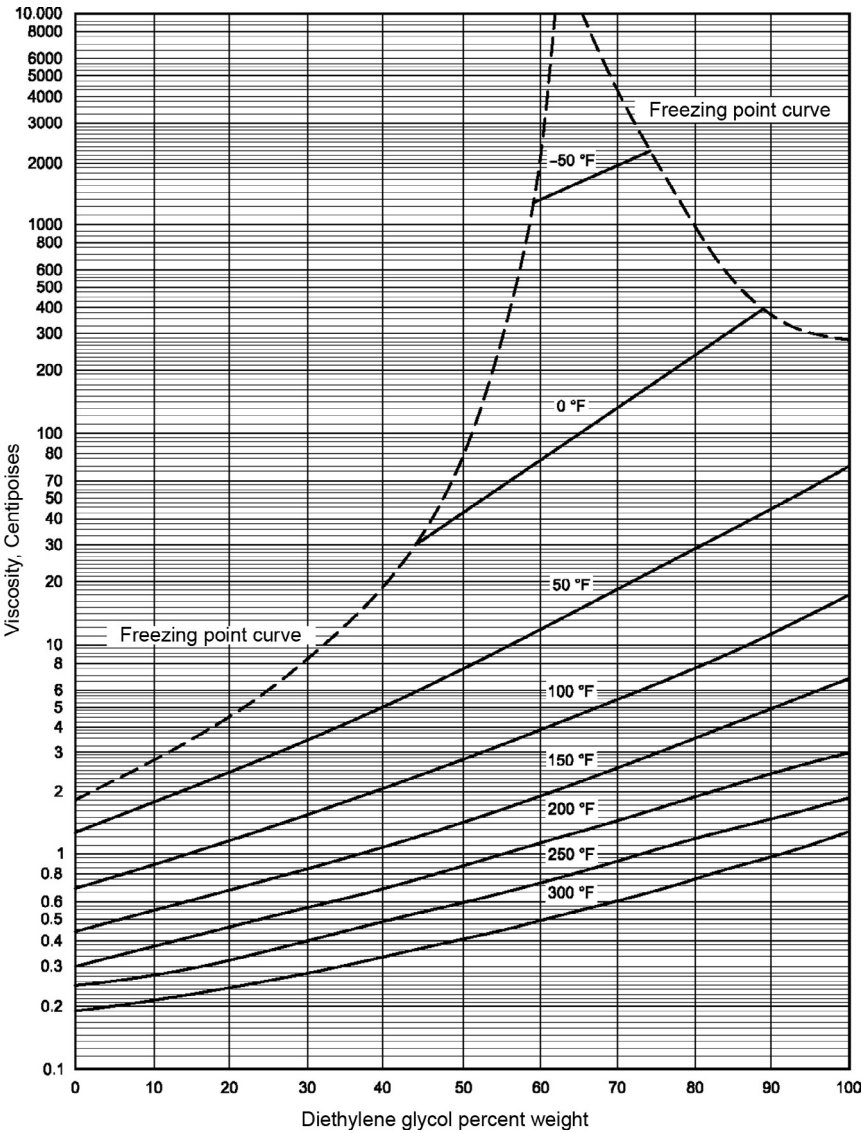


Figure 7.72 Viscosities of aqueous DEG solutions.

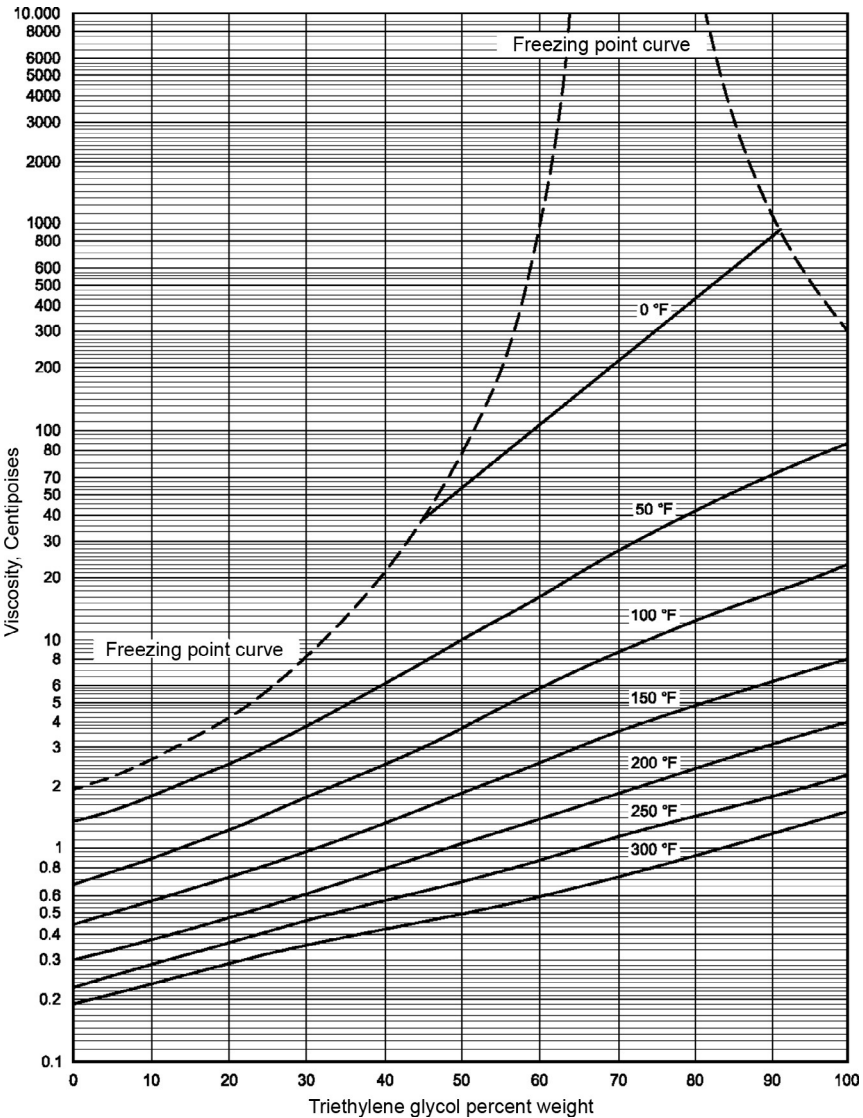


Figure 7.73 Viscosities of aqueous TEG solutions.



Glycol Maintenance, Care, and Troubleshooting



8.1 PREVENTIVE MAINTENANCE

8.1.1 Scheduled Preventive Maintenance

Scheduled preventative maintenance reduces glycol losses such as:

- Foaming
- System plugging

Maintenance also reduces mechanical failure such as:

- Corrosion
- Pump failures

It also minimizes system down time.

It maximizes system operation efficiency.

8.1.2 Five Steps to a Successful Preventive Maintenance Program

Record Keeping

Accurate records can be used to determine the system efficiency and to pinpoint operating problems.

Records of prior and existing conditions, including dew points, glycol usage, and repairs help establish the system profile.

Once the system profile is defined, it becomes easier to identify unusual system characteristics that may indicate potential problems.

Mechanical Maintenance

Daily physical inspections are necessary to ensure that the system is running properly.

Any trouble encountered should be dealt with immediately, thus preventing the problem from escalating.

Glycol Care

Regular chemical analysis (every 1 or 2 months) of the glycol provides detailed information about the internal operation of the unit.

Many process-related problems can be diagnosed well in advance of mechanical failure.

Chemical problems can be diagnosed and corrective action taken before they become costly and detrimental to the unit's performance.

Corrosion Control

Corrosion is a frequent problem in glycol dehydration systems (see [Figures 8.1–8.4](#)).

If unchecked, the damage from corrosion can be extensive.

All units should have provisions for corrosion control.

Communication

Lines of communication between field and office personnel are critical to the smooth operation of any system.

Office personnel (production supervisors, facility engineers, purchasing agents) must be kept informed of daily operations and any problems that may arise.

Field personnel must be made aware of technical information that may improve operations in the field.

Training for field operators allows the operator to better maintain the equipment.

8.1.2.1 Record Keeping

Records necessary to establish a system profile include design information, which includes vessel specifications, equipment drawings, and P&IDs

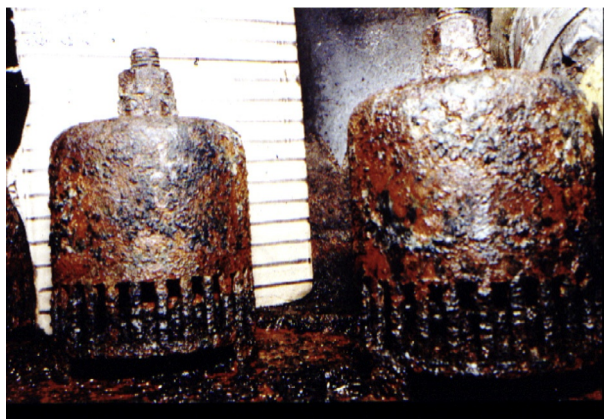


Figure 8.1 Oxidation corrosion of bubble cap tray.



Figure 8.2 Sweet corrosion of valve tray.

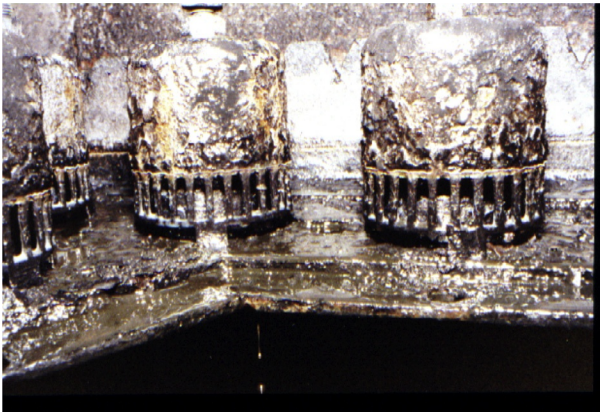


Figure 8.3 Sour corrosion of bubble caps.

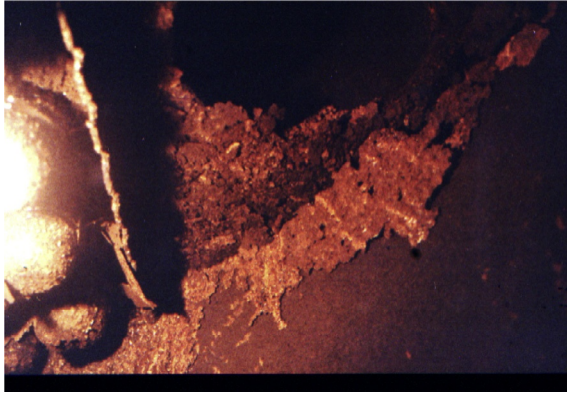


Figure 8.4 Generalized oxidation corrosion of contact tower.

Filter element or media replacement—type and frequency

Glycol usage—gallons/month

Chemical additives—type and amount

Gas production and flow rate charts—peak, average, and low periods

Outlet gas dew point/water content (lbs/MSMCF)

Mechanical inspections

Type

Magnitude

Recommendations

Results

Records necessary to establish a system profile include:

Glycol analysis—format, frequency, recommendations, results

Corrosion coupon results—mills per year (MPY), frequency

Materials and labor relating to system repairs—operating costs

Figures 8.5–8.7 are examples of common report formats.

With the aforementioned information, a good system profile can be drawn of a specific system.

Updating these records will show any gradual changes in a unit's system profile and may alert you to a potential or developing problem.

8.1.2.2 Mechanical Maintenance

The following things should be done to keep the unit operating properly and to prevent operational problems:

1. Ensure that instruments and controls are in good working condition (thermometers and pressure gauges, etc.). Use a test thermometer on the reconcentrator to ensure proper reconcentrator heat.

MONTHLY CHEMICAL USAGE

Company

Month & Year

Platform

Glycol Capacity

	Glycol		Corrosion Inhibitor		De-Foam	Inhibitor	
Day	Quantity Added in Gallons	Inventory	Quantity in Gallons	Inventory	Quantity in Gallons	Inventory	Comments
1							
2							
3							
4							
5							
6							
7							
8							
9							
10							
11							
12							
13							
14							
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21							
22							
23							
24							
25							
26							
27							
28							
29							
30							
31							

Figure 8.7 Monthly chemical usage.

2. Ensure glycol filter elements are changed according to average expected need:

Microfiber filters should be changed monthly.
Carbon filters should be changed monthly (small cartridge filters) to every 6 months (large bulk units).
Glycol analysis helps determine the frequency.
An upset or sudden change in the operating conditions may foul the filters faster than the preventive maintenance program anticipates.
Make sure filter differential pressure is below15 psi.
3. Look for glycol leaks on and around the glycol skid.

Most leaks can be stopped by tightening a union, valve stem packing, or pump rod packing.
After the leak has been repaired, clean the affected area so it is easier to notice new leaks.
4. Check the glycol level at least twice a day, and add glycol as necessary.

Maintain a written report of glycol added.
This allows operations to detect excessive losses of glycol and take corrective action faster.
5. Ensure unit performance by taking a dew-point measurement daily.
6. Clean the glycol strainers monthly to prevent accumulation of trash, which can cause the glycol pump to fail.

7. Check the glycol circulation rate daily.

Any time the gas flow rate changes, or when a drastic change in gas pressure or temperature is experienced, the glycol flow rate should be recalculated and the pumps set accordingly.

On multiple pump installations, switch the pumps weekly, thus ensuring pump operation when necessary.

8. Sight down the fire tube weekly on direct-fired firetubes for firetube blisters or hot spots.

These indicate extreme fouling and impending firetube failure.

9. Cycle the main burner manually to be sure the fuel gas valve works and the pilot light stays lit.

Check the fuel gas scrubber pot for fluid buildup that may hinder burner operation.

8.1.2.3 Glycol Care

8.1.2.3.1 General Considerations

Operating and corrosion problems occur when the circulating glycol gets dirty.

Some contaminated glycol problems can be noticed easily and corrective action taken.

A small glycol sample should be taken daily from the surge tank or the dry glycol suction header to the pump.

Check closely for fine black particles settling out of the sample, which may be corrosion byproducts and indicate an internal corrosion problem (see [Figures 8.8](#) and [8.9](#)).



Figure 8.8 Sample of uncontaminated glycol.

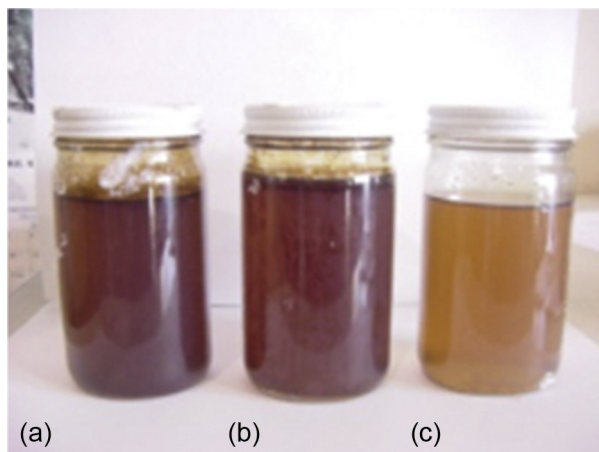


Figure 8.9 Three samples of liquid-hydrocarbon contaminated glycol: (a) moderate; (b) severe; (c) slight.

8.1.2.3.2 Smell the Sample

If the sample smells sweet and aromatic (similar to rotten bananas), it may be thermally decomposed.

If the sample is viscous and black, it is probably contaminated with hydrocarbon or well-treating chemicals.

If the hydrocarbon contamination is great enough, the sample will separate into two liquids or interphase.

Every 1 or 2 months, send a sample of both the rich and lean glycol to a laboratory for complete analysis.

This type of analysis will provide a detailed description of unit performance and glycol condition.

8.1.2.4 Corrosion Control

8.1.2.4.1 Overview

8.1.2.4.1.1 General Considerations

Corrosion is a major cause of premature equipment failure.

Corrosion can occur over the entire system, inside and out.

The two most common areas of severe corrosion are:

- Still column reflux coil

- Vent/fill connection on the surge tank

This is due to a high concentration of water vapor in the top of the still and the readily available oxygen in the air at the vent/fill cap.

Three types of corrosion that are almost always in glycol systems either individually or in combination with one another are:

Oxidation (Figure 8.1)

Sour corrosion (Figure 8.3)

Sweet corrosion (Figure 8.2)

8.1.2.4.1.2 Oxidation

Oxidation of metal is the exchange of electrons between metal and oxygen molecules to form positive and negative hydrogen ions.

Some metal loss is incurred.

The resultant scale-like residue formed by the process is called oxide, or rust.

Oxidization is characterized by rough, irregular, shallow pitting of the metal scaled over the rust.

8.1.2.4.1.3 Sour Corrosion

Acid gases (H_2S and CO_2) are often found in produced natural gas.

Glycols are very reactive with sulfur compounds, such as H_2S , and will exchange electrons with metal molecules, which initiate corrosion.

The resulting materials tend to polymerize (form larger molecules) that form a “gunk” that is very corrosive.

Sour corrosion is characterized by deep, jagged pitting.

8.1.2.4.1.4 Sweet Corrosion

Water is found in glycol as vapor, free condensed water, or entrained in glycol.

When carbon dioxide (CO_2) is dissolved in water it forms carbonic acid.

Because most produced natural gases contain some CO_2 , the presence of carbonic acid in glycol systems is very common.

The corrosion resulting from carbonic acid is known as sweet corrosion.

Sweet corrosion is characterized by deep, round, smooth pitting.

Sometimes the pitting will cover a broad area, disguising the depth of the pit.

8.1.2.4.2 Prevention and Control Programs

8.1.2.4.2.1 General Considerations

Prevention and control programs should include system monitoring through:

- Corrosion coupons

- Glycol analysis (pH and iron)

The three steps in combating corrosion in glycol systems are:

- Use an effective corrosion inhibitor in both the liquid and vapor phases.

- Use corrosion-resistant alloys in construction.

- Keep the unit clean to prevent acid formation due to contamination.

Cathodic protection has been attempted but has been met with little success.

It is impractical to attempt to eliminate corrosion completely.

The rate of corrosion can be slowed to a point that is almost negligible.

The maximum acceptable corrosion rate is 6 MPY.

Corrosion inhibitors work in several ways. The two most applicable to glycol units are:

- pH buffers

- Plating inhibitors

8.1.2.4.2.2 pH Buffers

pH buffers include:

- Alkanolamines

- MEA

- TEA

They fight corrosion by stabilizing the pH near neutral, thereby reducing the corrosive environment.

Amines are not true inhibitors in that no protection is actually afforded to the metal surface.

- Alkanolamines are regenerable, as is the glycol, and thus can be retained in the system for lengthy time periods.

- However, they are thermally degraded at normal reconcentrator operating temperatures and, if used frequently, may leave harmful residues within the system.

8.1.2.4.2.3 Plating Inhibitors

Tallow diamine, unlike inorganic amines, such as alkanolamines, is an organic amine.

It is grouped with the plating inhibitors even though it does not actually plate out on the vessel walls.

It flashes out of the glycol at high temperatures. As it vaporizes, it contacts the vapor spaces of the reconcentrator and forms a tenacious film over the exposed metal.

This film will eventually wear away and must be replenished occasionally to continue protection.

To a lesser degree than the alkanolamines, tallow diamine will also buffer pH.

True plating inhibitors include:

- Borax

- NaCap (Sodium mercaptobenzothiazole)

- Dipotassium phosphate

These inhibitors are strictly liquid phase protection.

They will plate out on the vessel walls forming a protective barrier between a corrosive environment and the metal.

This barrier also prevents the ready exchange of electrons, which drastically slows corrosion.

Since the plating inhibitors are all alkalies, some degree of pH buffering is effected.

The pH buffering is not as great as through the use of amines.

8.1.2.5 Communication

Communication is the easiest portion of an effective maintenance program and yet it is the most overlooked.

Communication can be between:

Management and labor

Engineer and foreman

Operators on opposite shifts

Office and field personnel

Lack of communication is the single most contributing factor to glycol system failure.

When were the glycol filters changed?

How much glycol losses are being experienced?

What are the results of glycol analysis?

What is the immediate history of the problem?

Failure to communicate can cause confusion and evolve into major problems.



8.2 GENERAL CONSIDERATIONS

Operating and corrosion problems usually occur when the circulating glycol gets dirty.

To achieve a long, trouble-free life from the glycol, it is necessary to recognize these problems and know how to prevent them.

Some of the major problem areas are:

Oxidation

Thermal decomposition pH control

Salt contamination

Hydrocarbons

Sludge

Foaming

8.2.1 Oxidation

Oxygen enters the system with the incoming gas through:

- Unblanketed storage tanks and sumps

- Pump packing glands

Sometimes glycol will oxidize in the presence of oxygen and form corrosive acids.

To prevent oxidation:

- Bulk storage tanks should be gas blanketed.

- Use oxidation inhibitors.

- Normally, a 50/50 blend of Monoethanolamine and 331/3% hydrazine is inserted into the glycol between the absorber and the reconcentrator.

Preferably, a metering pump should be used to give continuous, uniform injection.

8.2.2 Thermal Decomposition

Excessive heat, a result of the following conditions, will decompose glycol and form corrosive products:

- High reconcentrator temperature above the glycol decomposition level

- High heat-flux rate, sometimes used by a design engineer to keep the heater cost low

- Localized overheating, caused by deposits of salts or tarry products on the reconcentrator firetubes or by poor flame direction on the firetubes

8.2.3 pH Control

pH is a measure of the acidity or alkalinity of a fluid, based on a scale of 0-14.

- pH values from 0 to 7 indicate the fluid is acidic.

- pH values from 7 to 14 indicate the fluid is alkaline.

To obtain a true reading, glycol samples should be diluted 50-50 with distilled water before pH tests are run.

The pH meter should be calibrated occasionally to keep it accurate.

The pH of the still water should also be checked to ensure that it has the neutral value of 7.

New glycol has a neutral pH of ~7.

- With usage, the pH decreases and the glycol becomes acidic and corrosive unless pH neutralizers or buffers are used.

Equipment corrosion rate increases rapidly with a decrease in the glycol pH.

Acid created by glycol oxidization, thermal decomposition products, or acid gases picked up from the gas stream are the most troublesome of corrosive contaminants.

A low pH accelerates the decomposition of glycol. Ideally, the glycol pH should be held at a level between 7.0 and 7.5.

A value above 8.5 tends to make glycol foam and emulsify.

A value below 6.0 corresponds to system contamination, corrosion, and/or oxidation.

Borax, ethanalamines (usually triethanolamine) or other alkaline neutralizers are used to control the pH.

These neutralizers should be added slowly and continuously for the best results.

An overdose will usually precipitate a suspension of black sludge in the glycol.

The sludge could settle and plug the glycol flow in any part of the circulating system.

Frequent filter element changes should be made while pH neutralizers are added.

The amount of neutralizer to be added and the frequency will vary from location to location.

Normally, 1/4 lb of triethanolamine (TEA) per 100 gallons of glycol is sufficient to raise the pH level to a safe range.

When the glycol pH is extremely low, the required amount of neutralizer can be determined by titration.

For best results, the lean rather than the rich glycol should be treated. It takes time for the neutralizer to mix thoroughly with all of the glycol in the system.

Several days are required before the pH is raised to a safe level.

Each time neutralizer is added, the pH of the glycol should be measured several times.

8.2.4 Salt Contamination

8.2.4.1 Salt Deposits

Salt deposits accelerate equipment corrosion.

It also reduces heat transfer in the firetubes.

It alters specific gravity readings when a hydrometer is used to measure glycol water concentration.

It cannot be removed with normal regeneration. A scrubber installed upstream of the glycol plant should be used to prevent salt carryover from produced free water.

In areas where large quantities of brine are produced, some salt contamination will occur.

The removal of salt from the glycol solution is then necessary. The following reclaiming methods are used:

- Scraped surface heat exchangers in conjunction with centrifuges

- Vacuum distillation

- Ion exchange

- Ion retardation

8.2.5 Hydrocarbons

Liquid hydrocarbons, a result of carryover with the incoming gas or condensation in the contactor, increases glycol by:

- Foaming

- Degradation

- Losses

It must be removed with:

- Glycol/gas/condensate separator

- Hydrocarbon liquid skimmer

- Activated carbon beds

8.2.6 Sludge

Solid particles and tarry hydrocarbons (sludge) are suspended in the circulating glycol and with time will settle out (see [Figure 8.10](#)).

This action results in the formation of black, sticky, abrasive gum that can cause trouble in pumps, valves, and other equipment, usually when the glycol pH is low.

8.2.7 Foaming

8.2.7.1 General Considerations

Excessive turbulence and high liquid-to-vapor contacting velocities often cause the glycol to foam (this condition can be caused by mechanical or chemical problems).

The best way to prevent foaming is proper care of the glycol, such as:

- Effective gas cleaning ahead of the glycol system

- Good filtration of the circulating solution

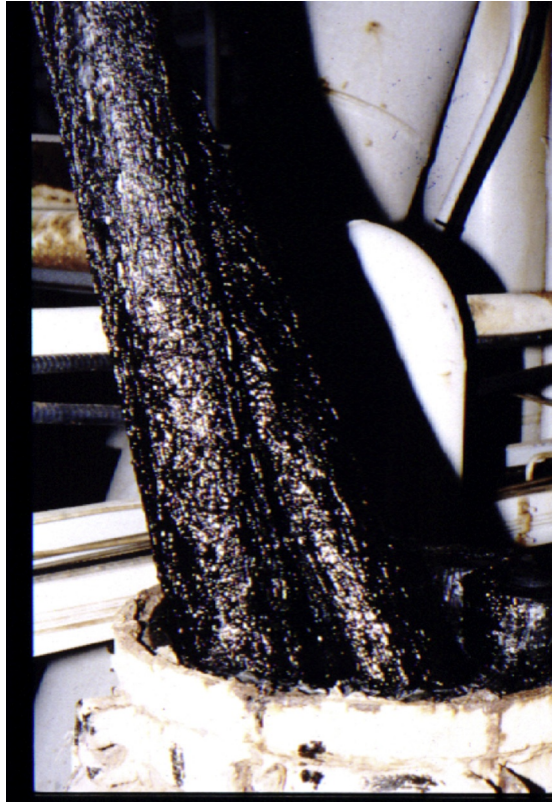


Figure 8.10 Sludge captured on glycol microfiber filters.

8.2.7.2 Defoamers

Defoamers serve only as a temporary control until the conditions generating foam can be identified and removed.

Success depends on when and how it is added.

Some act as good inhibitors if they are added after the foam has been generated, but aggravate the problem if they are added prior to the onset by serving to stabilize the foam.

Most are inactivated within a few hours under high temperature and pressure, and thus their effectiveness is dissipated by the heat of the glycol solution.

Defoamers should be added continuously, a drop at a time, for best results.

Chemical feed pumps:

Meter the defoamer accurately

Improve dispersion into the glycol

Are activated automatically by differential pressure across the contactor

8.2.8 Analysis and Control of Glycol

8.2.8.1 General Considerations

Analysis of glycol is essential to good plant operation.

It helps pinpoint high glycol losses, foaming, corrosion, and other operating problems.

Analyses enable operations personnel to evaluate plant performance and make operating changes to obtain maximum drying efficiency.

8.2.8.2 Visual Inspection

A glycol sample should first be visually inspected to identify some of the contaminants (see Figure 8.11).

A finely divided black precipitate may indicate the presence of iron corrosion products.

A black, viscous solution may contain heavy, tarry hydrocarbons.

The characteristic odor of decomposed glycol (a sweet aromatic odor) usually indicates thermal degradation.

A two-phase liquid sample usually indicates the glycol is heavily contaminated with hydrocarbons.

The visual conclusion should be supported by a chemical analysis.

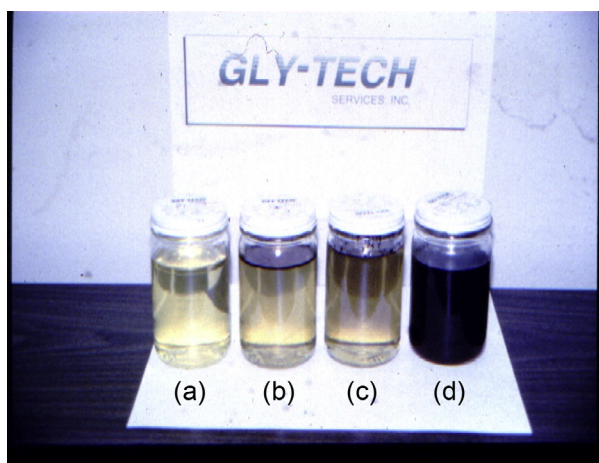


Figure 8.11 Glycol samples: (a) normal glycol; (b) hydrocarbon carryover; (c) iron corrosion particles settling to the bottom of sample container; (d) two-phase large liquid hydrocarbon carryover. *Courtesy of Gly-Tech.*

8.2.9 Chemical Analysis

A complete glycol analysis of lean and rich samples, when properly interpreted, can provide a detailed picture of the workings of the dehydration unit and its process.

Glycol analysis should include tests to determine the following (refer to Table 8.1):

- pH (50/50)
- Hydrocarbon content (% wt.)
- Water content (% wt.)
- Total suspended solids (% wt.)
- Residue (% wt.)
- Chlorides (mg/l)
- Iron (mg/l)
- Foam character
- Height (ml)
- Stability (s)
- Specific gravity

Table 8.1 Chemical Analysis

Company	Date:			
Location				
Test	Lean Glycol	Rich Glycol	Allowable Range	Ideal
pH (50/50)			6-8	7-7.5
Hydrocarbon (% wt.)			0.1%	
Water content (% wt.)			2% lean 6% rich	
TSS (% wt.)			0.01%	
Residue (% wt.)			4%	2%
Chlorides (mg/l)			1500	1000
Iron (mg/l)			50	35
Foam character				
Height (ml)			20-30 ml	
Stability (s)			15-5 s	
Specific gravity			1.118-1.126	
Glycol composition				
EG				
DEG				
TEG				
TTEG				

Glycol composition

EG

DEG

TEG

TTEG

8.2.10 Chemical Analysis Interpretation

8.2.10.1 pH

A pH below 6 generally corresponds with system contamination, corrosion, and/or oxidation.

Below 5.5 autoxidation occurs.

Chemical decomposition of the glycol occurs within itself.

Glycol has the tendency to continue to drop without outside influences.

Causes of low pH:

Acid gases in the gas stream

Organic acids due to oxidation or thermal degradation

Excessive chlorides (salt) in the glycol

Well-treating chemicals entrained in the gas stream

Thermal decomposition of entrained liquid hydrocarbons in the gas stream and glycol

Oxidation of the glycol due to improper storage

Causes of high pH:

Contamination from well-treating chemicals entrained in the gas stream

Overdose of neutralizer added to a system for low pH

Foaming tendencies can result from high pH, due to stabilized glycol-hydrocarbon emulsions.

Sludge and residue buildup can result from both high and low pH.

8.2.10.1.1 Sludge

May become abrasive and cause premature pump and valve failure.

May deposit in trays and downcomers, still column packing, and heat exchangers, which causes system plugging.

8.2.10.1.2 Hydrocarbons

Hydrocarbons enter the glycol stream as a result of inlet separator carry-over or as condensation due to temperature variations.

Compressor lube oils and other extraneous organic chemicals, such as pipeline corrosion inhibitors, are often stripped out of natural gas as it passes through the contact tower.

Oils and organic residues can cause glycol/water emulsions and suspensions, which contribute to foaming:

- Results in excessively high glycol carryover from the contactor

- Contaminants may cause plugging in the contactor, still column, and heat exchangers

Light hydrocarbons:

- Usually separated from the glycol stream with an adequately sized glycol/ hydrocarbon separator

Heavy hydrocarbons:

- Referred to as soluble hydrocarbons because they bond with the glycol

- Usually filtered out with activated carbon

Light end hydrocarbons (insoluble) are allowable up to 1% by volume.

Soluble hydrocarbons are only acceptable to 0.1% by weight.

- Primarily responsible for foaming, sludge, and residue buildup, low pH, loss of hygroscopicity, and glycol decomposition.

8.2.10.1.3 Water Content

Water content is defined as the quantity of water in the glycol.

The difference between the lean sample and rich sample measures the degree of loading in the contactor.

It indicates regeneration efficiency.

Glycol purity should be at least 98% in the lean stream and at least 94% in the rich.

- These concentrations will produce the desired dew points in systems that are operating properly.

- For lower dew points, the glycol purity must be increased (or water content decreased).

High water content of the lean sample generally indicates low reconcentrator heat.

High water content in the lean sample may also indicate:

- Excessive glycol circulation

- Undersized equipment

- Carryover from the separator

- Vapor communication from reconcentrator to surge

- A leak in the glycol/glycol heat exchanger

- Over-refluxing in the still column
- Hot inlet gas temperature

High water content in the rich sample usually indicates a low glycol circulation rate or:

- Carryover from the separator
- Poor reconcentration
- Heat exchanger communication
- Undersized equipment
- Hot inlet gas temperature

Check values for hydrocarbon, chlorides, iron, and foaming to help pinpoint the problem.

8.2.10.1.4 Suspended Solids

Suspended solids are considered to be those solids and tarry hydrocarbons that remain suspended within the glycol solution down to $0.45\text{ }\mu\text{m}$ in size.

They are the result of poor inlet separation, corrosion, and thermal degradation of the glycol.

Values $>0.01\%$ by weight indicate poor sock/microfiber filtration.

Most filters are sized to remove particles to a size of $5\text{ }\mu\text{m}$.

Particles larger than this in excessive amounts may serve to stabilize foaming tendencies in glycol.

When the glycol is allowed to maintain a large concentration of suspended solids, a silty residue is likely to form along vessel walls.

Plugging of the contractor trays, heat exchangers, still column, and reconcentrator glycol is likely (common with low glycol pH).

8.2.10.1.5 Residue

The value for residue is a function of system contamination.

The glycol sample is distilled, removing all light end hydrocarbons, water, and glycol.

Residue represents the remaining contamination, which is comprised of:

- Total solids (suspended and residual)

- Salt

- Heavy hydrocarbons

Value for residue is best kept below 2% by weight, however, some systems may operate reasonably well at values from 2% to 4%.

Units with glycol containing $>4\%$ are prime candidates for failure and should be cleaned immediately.

8.2.10.1.6 Chlorides

Chloride values indicate the quantity of inorganic chlorides (salts) found in the glycol sample.

As the concentration of chlorides (as NaCl or CaCl) in glycol increases, its solubility decreases.

Solubility also decreases when heat is added to the glycol solution.

When the solubility decreases, the salt begins to form crystals which:

- Fall out of the glycol solution

- Accumulate on the heat source and can lead to premature heat tube failure

- May be swept by the glycol into other areas of the system

Potential problems with excessive chlorides include system plugging, low pH, glycol pump damage, foaming, and loss of hygroscopicity due to rapid glycol decomposition.

Removal of chlorides in high concentrations requires vacuum distillation of the glycol.

Concentrations >1000 ppm will stabilize foaming tendencies.

- May lead to excessive glycol loss

- May affect glycol pH

Precipitation of salts from the glycol will begin at ~12,200–1500 ppm, however, the crystals formed are extremely small and rarely troublesome. Concentrations above 2200 ppm precipitation occurs readily, and system failure is a possibility.

Filtration removes large salt crystals, but most of the damage associated with salt will have already occurred prior to the development of crystals large enough to filter.

8.2.10.1.7 Iron

Iron found in glycol samples can indicate:

- Possible corrosion

- Produced water carryover

Iron in excess of 50 ppm generally indicates corrosion.

Whether it be in the glycol unit, upstream in the production equipment, or downhole in the well string, it is difficult to determine.

Comparing values for pH, chlorides, and visual inspection of the glycol unit may help to establish the location of suspected corrosion.

Corrosion by-products will consist of soluble iron and fine, gritty particulate in systems where oxygen is available.

In systems where no oxygen is present, corrosion by-products will include sulfides in addition to the iron.

8.2.10.1.8 Foaming

General Considerations

More glycol is lost through foaming than by any other cause.

It is not easily detected without chemical analysis; gradual low-volume glycol loss often goes overlooked.

It is almost always a result of contamination. Primary contaminants that cause foaming are:

Hydrocarbons (from separator carryover) and solids

Chlorides, compressor lube oil, well-treating chemicals, and iron

Water content affects foaming tendencies by inducing emulsification of contaminants, particularly hydrocarbons.

Carbon filtration is the most effective means of controlling foam.

Silicone emulsion-type foam inhibitor may be used, but they treat the symptoms, not the cause, and are temporary solutions.

Addressing the source of the contamination causing the foam is the only long-term solution.

Foam Test

The foam test consists of bubbling dry air at a rate of 6 liters/min through a graduated cylinder container of 200 mm of the glycol sample until the foam stabilizes at its maximum height.

Volume for both the liquid and the foam is reported as a single value. The original 200 ml is then subtracted, and the remaining value is recorded as height and represents the ease at which the solution will foam.

Once the maximum foam height is recorded, the dry air is removed from the sample and the time it takes for the foam to break from its maximum volume to a clear surface on the glycol sample is recorded in seconds.

This time represents the tendency of the foam and is known as stability. There are no concrete values given for acceptable foam height and stability.

Foam with very low height and moderate stability will result in little glycol loss as will foam with moderate height and very low stability.

Thus, the acceptable range for foam test results are:

Height/ml: 20-30 ml

Stability/sec: 15-5 s

Acceptable Limits

Acceptable limits for foam represent the increasing and decreasing values of acceptability for height and stability.

For example, a sample with a height of 25 ml and a stability of 10 s is acceptable, while a sample with 30 ml height and 15 s stability would have a high foaming tendency and could result in glycol losses.

8.2.10.1.9 Specific Gravity

Specific gravity is used to determine the purity of glycol.

A specific gravity of 1.126–1.128 at 60 °F indicates a 99% TEG (technical grade).

A specific gravity of 1.124–1.126 indicates 97% (industrial grade).

With glycol extracted from an operating dehydration unit, the lean sample should have a specific gravity of 1.1189–1.121.

This variance allows for acceptable amounts of system contamination.

Low specific gravity would indicate one or more of the following:

- Triethylene Glycol containing excessive amounts of Ethylene Glycol and/or Diethylene Glycol (poor quality replacement glycol)

- Excessive water in sample

- Excessive hydrocarbons in sample

A high specific gravity indicates the system is contaminated with excessive amounts of solids or any additives with a greater density than glycol:

- Thermal degradation of the glycol

- Oxidation or chemical degradation of the glycol

8.2.10.1.10 Glycol Composition

The composition of glycol indicates its quality.

Values are given to the component glycols (EG, DEG, TEG, TTEG) contained within the glycol sample solution.

Industrial grade (97%) TEG or better is required for the best glycol system results.

In addition to 97% TEG, the glycol solution may contain, in various concentrations, up to 1% EG and 3% DEG, but is not to exceed a combined total of 3%.

Glycol degradation will often be reflected by changes in the glycol composition and reduction in pH.

Thermal degradation is most common and is characterized by excessive values of EG, DEG, and occasionally the presence of Tetraethylene (TTEG).

The glycol pH will be low.

The glycol sample will be dark and have an aromatic smell (ripe bananas). Chemical degradation is brought about by oxidation and acidic contaminants and is characterized by:

Excessive values for EG and DEG, but no TTEG, will be present Low pH. Glycol may not appear to be too dirty.

Autoxidation is a form of continuing chemical degradation.

8.2.11 Troubleshooting

8.2.11.1 General Considerations

Even the best preventive maintenance program will not guarantee that the dehydration unit will operate trouble-free.

The most obvious indication of a unit malfunction is high water content (dew point) of the outlet stream.

High water content is brought about by:

- Insufficient glycol circulation

- Reconcentration of the glycol

These problems can be caused by a variety of contributing factors such as:

- Mechanical causes

- Existing operating conditions for which the equipment was not designed

These conditions can sometimes be at least partially alleviated by changes in both condition and mechanical operation.

8.2.11.2 High Dew Points

8.2.11.2.1 Insufficient Glycol Circulation

If there is insufficient glycol circulation, check heat exchangers and glycol piping for restrictions or plugging.

8.2.11.2.1.1 On an Electricity-Driven Piston Pump

Check the flow rate indicator (if present) to ensure proper glycol circulation. If a flow rate indicator is not present, verify the circulation rate by closing the glycol discharge valve from the contactor and timing the fill rate in the gauge column.

Check the high-pressure dry-glycol bypass valve, and close it, if necessary.

Check the pump prime by shutting the pump down, closing the discharge valve, opening the bypass valve, and restarting the pump. Allow

it to briefly run under no load through the bypass line to remove any trapped gas in the pump.

8.2.11.2.1.2 On a Glycol-Gas Powered Pump

Close the dry discharge valve. If the pump continues to run, open the dry discharge bleed valve and allow it to run a few strokes. Once all of the gas is purged from put, close the bleed valve. If the pump continues to run, discontinue use and send it for repair.

If the pump will not prime, but continues to run gas through the dry discharge, bleed the valve then:

Check the pump suction strainer for plugging.

Check the glycol level in surge tank.

8.2.11.2.2 Insufficient Reconcentration

Verify the reconcentration temperature with test thermometer (350–400 °F), and raise the temperature if necessary.

Check the glycol-to-glycol heat exchanger for leakage of wet glycol into the dry glycol stream.

Check the stripping gas, if applicable. Be sure the stripping gas is in service at the proper rate.

Check for communication between the reconcentration vapor space and the surge tank vapor space. Communication could mean that contaminated dry glycol is going to the pump.

8.2.11.2.3 Operating Conditions Different from Design

Check the operation of upstream separators and scrubbers, and be sure not to overload the system.

Increase the absorber pressure; this may require installation of a back-pressure valve.

Reduce the gas temperature, if possible. Increase circulation rate, if possible. Increase reconcentrator temperature, if possible.

8.2.11.2.4 Low Flow Rate

Blank off a portion of the bubble caps, if possible.

Lower the system pressure.

Add additional cooling to the dry glycol and increase the circulation rate.

Change out the absorber to a small unit designed for a lower flow rate.

8.2.11.2.5 Absorber Tray Damage

Open inspection ports and/or manways and verify the tray integrity. Repair or replace the trays, as necessary.

8.2.11.2.6 Breakdown or Contamination of Glycol

Have both lean and rich glycol samples analyzed. Note any evidence of severe contamination—thermal or chemical decomposition. Clean the system and/or recharge it with fresh glycol as necessary.

8.2.11.3 *Glycol Loss from the Contactor*

8.2.11.3.1 Foaming

The major cause of foaming is contamination, so the source of contamination should be removed. Clean the contactor, and, if necessary, clean the system and replace the glycol.

Increase filter capacity and/or add carbon filtration.

Add antifoam compound (silicon emulsion type).

Adjust high pH to prevent emulsification (use acetic acid).

8.2.11.3.2 Plugged or Dirty Trays

Clean the trays.

Manually enter the tower and clean.

Open inspection ports and clean them with water jet or by hand.

Chemically clean.

8.2.11.3.3 Excessive Velocity

Decrease the gas rate.

Increase the absorber pressure.

8.2.11.3.4 Interrupted Liquid Seal on the Trays (Gas Surge)

If the contactor has a bypass valve, isolate the tower by opening the bypass valve and close the gas inlet valve. Allow the glycol pump to run 5 min. While the glycol is circulating, open the gas inlet valve and slowly close the gas bypass valve.

If the contactor does not have a bypass valve, stop or greatly reduce the gas flow through the tower (shut wells, flare gas, alternate system, etc.). Allow the glycol to circulate 5 min then slowly turn the gas back through the tower.

If unable to stop or reduce the gas flow, increase the glycol circulation rate to the maximum possible for 2-5 min (flood the trays in an attempt to reestablish a seal using liquid head pressure).

8.2.11.3.5 Cold Glycol (Cold Gas)

Increase the gas temperature by increasing the temperature of the flowline heater, or add a flowline heater, if necessary.

8.2.11.3.6 Leaks

Pressure test the external gas-glycol heat exchanger for glycol leakage into dry gas stream.

Check the drain header (if applicable) at all gauge columns, external float cages (LSLL, etc.).

8.2.11.3.7 Accumulation in Integral Scrubber

Check for communication between the chimney tray and the scrubber section.

Check the bottom tray for leakage. It may have a damaged or improperly constructed chimney.

Check the glycol level control and dump valve operation (on units with electricity-powered glycol pumps).

8.2.12 Glycol Loss from the Reconcentrator

8.2.12.1 Leaks

Be sure all drain valves are closed.

Be sure gauge column seals are in good repair.

Check heat tube integrity (glycol loss into fire tube or waste heat tube will produce heavy smoke from stack).

Check reconcentrator shell integrity (note glycol leakage from insulation, wet insulation, or telltale stains).

Heat source flange leak (poor gasketing).

8.2.12.2 Bad Glycol Relief Valve

Replace the glycol relief valve.

8.2.12.3 Exiting the Still Column

For plugged or fouled still column packing, clean or replace the still column packing.

For saturated glycol (droplets blowing out still): Check the reconcentrator heat source, and ensure that heat is between 350 and 400 °F. Check for free liquid or misting liquid carryover into the contactor tower. Repair or replace the separator control, if necessary. Reduce slugging if possible, and add a scrubber, if necessary.

Reduce glycol flow through the reflux condenser. Raise the reflux temperature.

8.2.12.4 Vaporization

Check the reconcentrator temperature (below 404 °F).

Check reflux temperature, and increase the glycol flow through the reflux condenser to lower the reflux temperature.

Check stripping gas flow rate.

Check for plugged or fouled glycol outlet from the reconcentrator (downcomer or heat exchangers).

8.2.13 Glycol Loss-Glycol Hydrocarbon Separator

8.2.13.1 Improper Control Operation

Repair or replace the level control.

Clean, repair, or replace the dump valve.

8.2.13.2 Leaks

Check the drain valve, and tighten, repair, or replace it, as necessary.

Check gauge columns, external float cages, and level control adapters.

Add antifoam compound to prevent loss through gas outlet.

8.2.13.3 Accumulation in Oil Bucket (Bucket-and-Weir)

Open the vessel and clean the glycol passage under the oil bucket (horizontal vessels).

Adjust or remove the weir.

8.2.14 Glycol Loss-Miscellaneous

8.2.14.1 Leaks

Check all flanges, unions, and associated piping.

Check electric pump rod packing.

Check all drain valves (filter, heat exchanger, etc.). Check the pump bleed valves (and electric pump bypass).

Check the external gas-glycol heat exchanger.

8.2.14.2 Poor Quality or Contamination Replacement Glycol

Use only virgin TEG with 97% or better purity.

Check glycol content for excessive water.

8.2.15 Three-Step Approach to Troubleshooting

8.2.15.1 Timeframe

Determine the approximate date/time at which the problem became apparent.

8.2.15.2 List Changes

Inventory any changes (things that happened differently than usual), and look for what is different.

Production changes

Operational changes

Maintenance

Repairs

Weather

8.2.15.3 Investigate

By process of elimination, reduce the list of changes to determine the factor or factors that manifest the problem.

8.2.16 Glycol System Cleaning

8.2.16.1 General Considerations

Chemicals are frequently needed to clean the glycol system.

If chemical cleaning is properly done, it can be quite beneficial to plant operations.

If chemical cleaning is poorly done, it can be quite costly and create long-lasting problems.

The most effective type of cleaner is a very heavy-duty alkaline solution. To provide optimum cleaning, the concentration, temperature, and pumping rate of the solution must be carefully controlled and an experienced, reputable vendor employed.

A cascading technique can be used to save on the cost of cleaning chemicals.

8.2.16.2 Cleaning Techniques to Avoid

Steam cleaning is not effective and can be damaging and dangerous. It tends to harden the deposits in the system, making them almost impossible to remove.

The use of cold or hot water, with or without high detergent soaps, will do little good in cleaning the system.

High-detergent soaps can create a serious problem by leaving trace quantities of soap behind.

Soap traces left in the system can create glycol foam for a long time.

Acid cleaning is good for removing inorganic deposits.

Because most deposits in the glycol system are organic, acid cleaning is not very effective.

Acid cleaning can easily create additional problems in the glycol system after the cleaning job.



8.3 ELIMINATING OPERATING PROBLEMS

8.3.1 General Considerations

Most operating problems are caused by mechanical failure.

It is important to keep equipment in good working order.

Following operating and maintenance suggestions helps provide a trouble-free operation.

8.3.2 Inlet Scrubber/Microfiber Filter Separator

The cleaner the inlet gas entering the absorber, the fewer operating problems there will be.

Potential problems when there is no inlet scrubber or filter separator are:

- Liquid water carryover

- Diluted glycol

- Lowered absorber efficiency

- A greater glycol circulation rate is required

- Increased vapor-liquid load on the still column

- Flooded still column

- Vastly increased reboiler heat load and fuel gas requirements

If the water contains salt and solids, they would be deposited in the reboiler to foul the heating surfaces and could possibly cause them to burn out (Figures 8.12–8.14)

If liquid hydrocarbons are present:

- They will pass onto the still column and reboiler

- Lighter fractions will pass overhead as vapor and could create a fire hazard

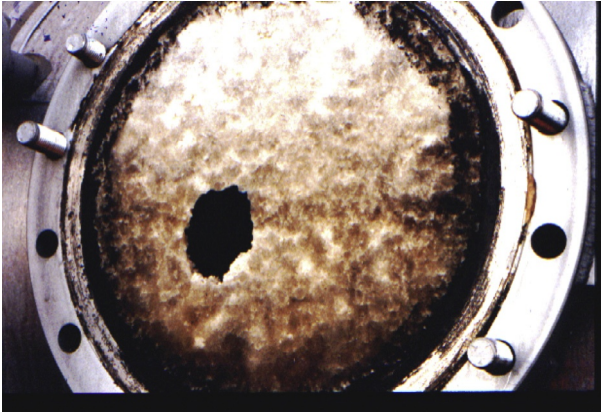


Figure 8.12 Salt deposited in interior of reboiler. *Courtesy of Gly-Tech.*



Figure 8.13 Salt fouling firetube. *Courtesy of Gly-Tech.*

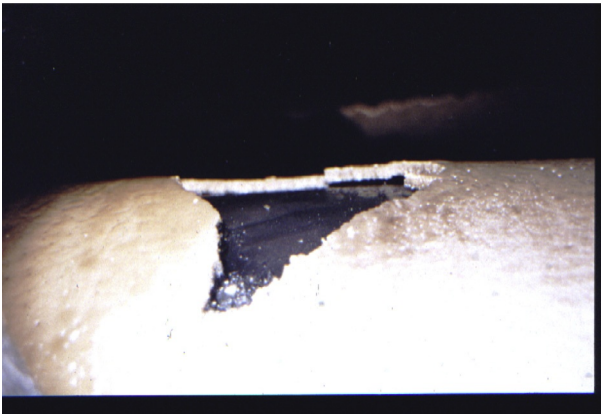


Figure 8.14 Salt covering firetube perimeter. *Courtesy of Gly-Tech.*

Heavy fractions will collect on the glycol surface in the storage tank and could overflow the system (Figure 8.15)

Flashing of the hydrocarbon vapor can flood the still column and greatly increase the reboiler heat load, resulting in glycol losses

A well corrosion-control program should be planned and coordinated to prevent glycol contamination:

Excessive fluid will carry over into the system if the inlet scrubber/filter separator is overloaded.

Gas from the treated wells should be slowly passed through a tank or separate system at the wellhead until the corrosion inhibitor and distillate carrier can be collected.

Do not open all of the treated wells at one time. This will keep large slugs out of the gathering lines going to the plant.

The scrubber or filter separator may be an integral part of the absorber or preferably, a separate vessel

The vessel should be large enough to remove all solids and free liquids to keep impurities from getting into the glycol system

The vessel should be regularly inspected to prevent any malfunction

The liquid dump line should be protected from freezing during cold weather:

Using a heating coil in the scrubber or separator

Pumping warm glycol through the coil

Directing flow through the coil by means of block and bypass valves

Provided a heating chamber on the liquid level controller and in the gage glass



Figure 8.15 Heavy fractions of tarry hydrocarbons collected from bottom of reboiler.

Including a heating coil in the reboiler to heat purge gas, which can be bled into the separator liquid dump line to keep the liquid moving so it does not freeze

The separator should be located close to the absorber so that the gas does not condense more liquids before it enters the absorber

If a separator ahead of the glycol plant is equipped with a pressure safety valve, a flow safety valve should be installed at the inlet to the absorber to protect the internals

Sometimes, an efficient mist extractor, which removes all contaminants over 1 μm , is needed between the inlet separator and the glycol plant to clean the incoming gas; this is particularly useful when paraffin and other impurities are present in a fine vapor form

When gas is compressed prior to dehydration:

A coalescing type of scrubber (microfiber filter separator) placed ahead of the absorber ensures removal of compressor oil in vapor form.

Compressor oil and distillate can coat the tower packing either in the absorber or the still column and decrease its effectiveness.

8.3.3 Absorber

The absorber is a vessel that contains valve or bubble cap trays or packing to give good gas-liquid contact.

Cleanliness is very important to prevent high sales gas dew points caused by foaming and/or poor gas-liquid contact.

Plugged trays or packing could also increase glycol losses.

Unit startup considerations are as follows:

The pressure on the absorber should be slowly brought up to the operating range, and then the glycol should be circulated to get a liquid level on all trays.

Next, the gas rate going to the absorber should be slowly increased until the operating level is reached.

If the gas enters the absorber before the trays are sealed with liquid, it will pass through the downcomers and bubble caps.

When this condition occurs and the glycol is pumped into the absorber, the liquid has difficulty sealing the downcomers.

Liquid will be carried out with the gas stream instead of flowing to the bottom of the absorber.

Gas flow rate should be increased slowly when changing from a low to a high flow rate.

Rapid surges of gas through the absorber may cause:

- Sufficient pressure drop through the trays to break the liquid seals, and/or

- Glycol to be lifted off the trays, which will flood the mist extractor and increase glycol losses

Unit shutdown considerations are as follows:

- First, the fuel to the reboiler should be shut down. Then, the circulating pump should be run until the reboiler temperature is lowered to $\sim 200^{\circ}\text{F}$ (94°C).

- This precaution will prevent glycol decomposition caused by overheating.

- The unit can then be shut down by slowly reducing the gas flow to prevent any unnecessary shocks on the absorber and piping.

- The unit should be depressurized slowly to prevent a loss of glycol.

- The dehydrator should always be depressurized from the downstream (gas outlet) side of the absorber.

A dehydrator installed on the discharge side of a compressor should be equipped with a check valve in the inlet line, located as close as possible to the absorber.

- Experience has shown that some glycol is sucked back into this line when a compressor backfires or is shut down.

- Internal absorber damage to the trays and mesh pad may also occur with a compressor failure (Figure 8.16).

- The installation of the check valve usually eliminates this problem.

All compressors taking gas from or feeding gas to a dehydrator should have pulsation dampeners.

- The absence of this safety device may cause fatigue failure of instruments, trays, coils, mesh pads, and other parts of the dehydrator.

The glycol dump valve and level controller should be set for throttling action to give an even flow of glycol to the regenerator.

- This will prevent slugs, which could flood the stripper and cause excessive glycol losses.

The absorber must be vertical.

- Ensure the proper flow of glycol in the vessel and adequate contact of the glycol and gas.

- Sometimes the trays and bubble caps do not seal properly after erection and should be inspected if very high glycol losses exist.



Figure 8.16 Tray damage due to compressor failure.

Inspection ports at the trays can be very useful when inspecting or cleaning the vessel.

If dry gas from a glycol unit is used for gas lift:

Care must be used in both sizing and operating the unit because of the unsteady gas rate required in this service.

A back-pressure valve should be installed on the gas outlet from the absorber operating on a gas lift system.

If this is not done, then a valve downstream of the absorber can be pinched to prevent a sudden overloading of the absorber and help control the gas flow through the unit.

A sudden overloading of the absorber can break the downcomer seals in a tray type of vessel and cause excessive loss of glycol in the sales gas.

Absorbers may need to be insulated when excessive condensation of light hydrocarbons collect on the vessel walls.

This often occurs when dehydrating rich, warm gases in cold climates.

These very light hydrocarbons can cause tray flooding in the absorber and excessive glycol losses from the regenerator.

The mist extractor should receive special attention because glycol entrainment and well-crawling are difficult to effectively control (see [Figures 8.17–8.20](#)).

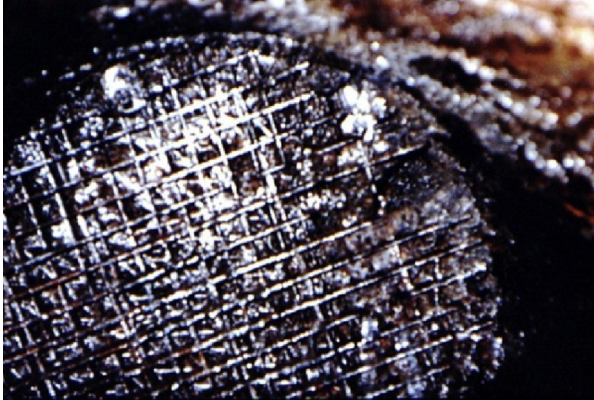


Figure 8.17 Partially plugged mist extractor.

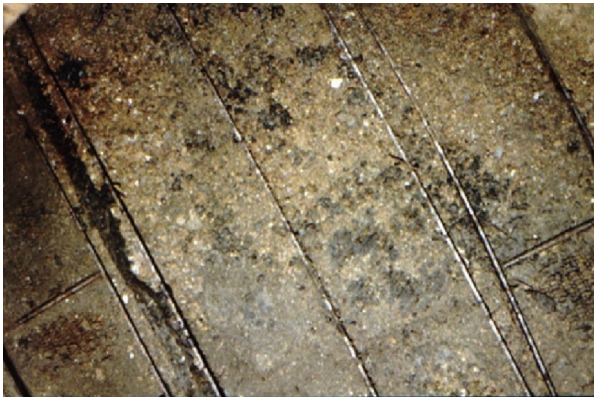


Figure 8.18 Completely plugged mist extractor.



Figure 8.19 Replacement mist extractor. *Courtesy of Gly-Tech.*



Figure 8.20 Installation of replacement mist extractor.

The type and thickness of the mesh pad should be carefully studied to minimize glycol losses.

Care should also be taken after installation to avoid mesh pad damage.

The maximum pressure drop through the contractor to avoid damage to the mesh pad is ~ 15 psi.

8.3.4 Glycol-Gas Heat Exchanger

Most units are supplied with a glycol-gas heat exchanger that uses the gas leaving the absorber to cool the lean glycol entering the absorber.

This exchanger may be a coil in the top of the absorber or an external one.

A water-cooled exchanger may be used when heating of the gas must be avoided.

This exchanger may accumulate deposits, such as:

- Salt
- Solids
- Coke
- Gum

Which

- Foul the heat exchanger surface
- Reduce the heat transfer rate
- Increase the lean glycol temperature.

All of these actions increase glycol losses and make dehydration difficult.

The vessel should be inspected regularly and cleaned when needed.

8.3.4.1 Lean Glycol Storage Tank or Accumulator

Normally, this type of vessel contains a glycol heat exchanger coil, which does the following:

- Cools the lean glycol coming from the reboiler

- Preheats the rich glycol going to the stripper

The lean glycol is also cooled by radiation from the shell of the storage tank.

This accumulator should normally be insulated.

Water cooling can also be used to help control the lean glycol temperature.

On conventional regenerators without stripping gas:

- Accumulators must be vented to prevent trapping gas

- Vapors, trapped in the storage tank, could cause the pump to vapor lock

- A connection is usually provided in the top of the storage tank for venting

- Vent lines should be piped away from the process equipment but should not be connected to the stripper vent because this could cause steam to dilute the concentrated glycol

Some units are equipped to provide a dry gas blanket (no oxygen or air) in the storage tank.

- It is not usually necessary to attach a separate vent on these storage tanks.

- Blanket gas is normally piped to the regular vent connection on top of the storage tank.

- If blanket gas is used, it is commonly taken from the fuel gas line.

- When blanket gas is used, it may be necessary to see that the blanket gas valve, piping, and flow control orifice are open to pass gas.

- Only a very slight flow of gas is required to prevent steam generated in the reboiler from contaminating the regenerated glycol.

- The vessel should be inspected occasionally to see that sludge deposits and heavy hydrocarbons are not collecting in the bottom of the vessel.

- The heat exchanger coil should be kept clean so that proper heat transfer can be made. Keeping the coil clean also prevents corrosion.

- If the heat exchanger develops a leak, the water-rich glycol could dilute the lean glycol.

Glycol level in the storage tank should be checked and a level in the gauge glass should always be maintained.

Gauge glass should be kept clean to ensure an optimum level.

Glycol should be added as the level is pumped down.

Records of the amount of glycol added should be maintained.

Make certain the storage tank is not overfilled as this could present problems as well.

8.3.4.2 Stripper or Still Column

The stripper, or still column, is generally a packed column located on top of the reboiler to separate the water and glycol by fractional distillation.

Packing is usually a ceramic saddle, however, 304 stainless steel pall rings can be used to prevent breakage (see [Figures 8.21–8.24](#)).

A standard stripper usually has a finned atmospheric condenser in the top to cool the steam vapors and recover the entrained glycol.

Atmospheric condenser depends on air circulation to cool the hot vapors.

Increased glycol losses can occur on extremely hot days when insufficient cooling in the condenser causes poor condensation.

High glycol losses can also occur on extremely cold, windy days when excessive condensation (water and glycol) overloads the reboiler.

Excess liquids percolate out of the stripper vent.

If stripping gas is used, an internal reflux coil is typically provided to cool the vapors.



Figure 8.21 Ceramic saddle packing.



Figure 8.22 Stainless steel pall ring packing.

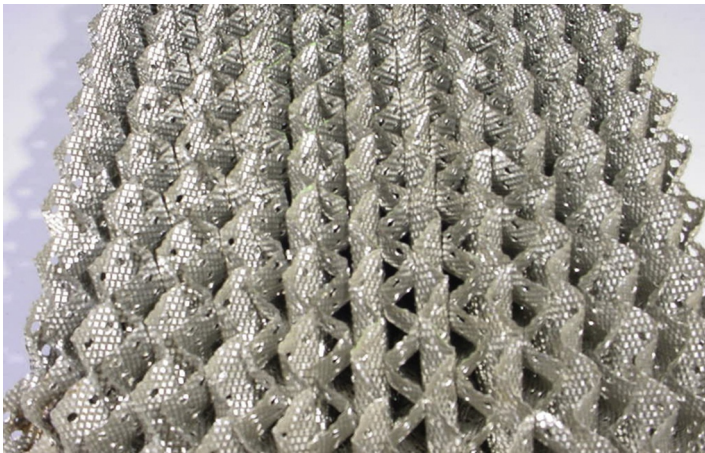


Figure 8.23 Structured packing.

Reflux for the stripper is more critical when stripping gas is used to prevent excessive glycol losses.

This is due to a larger mass of vapor leaving the stripper, which will carry glycol.

Adequate reflux is provided by passing the cool, rich glycol from the absorber through the condenser coil in the stripper.

If properly adjusted, it can provide uniform condensation throughout the year.



Figure 8.24 Ceramic saddles coated with hydrocarbon tar.

A manual/automatic valve in the piping is furnished to bypass the reflux coil.

Under normal circumstances, this valve is closed and the total flow is through the reflux coil.

In cold weather operation, with extreme low ambient temperatures, this could produce too much reflux and the regenerator could become overloaded.

The reboiler may not be able to maintain the required temperature. With these conditions, the ambient air is providing part or all of the reflux required.

Therefore, a portion or all of the rich glycol solution should bypass the reflux coil.

This is accomplished by opening the manual/ automatic valve until the reboiler can hold the temperature.

This lowers the amount of reflux produced by the coil and reduces the load on the reboiler.

Sometimes a leak can develop in the cool glycol reflux coil in the top of the stripper. When this happens, excess glycol can:

Flood the tower packing in the still column

Upset the distillation operation

Increase glycol losses

The reflux coil should be properly maintained.

Broken, powdered packing can cause solution foaming in the stripper and increase glycol losses.

Packing is usually broken by excessive bed movement, which is caused when hydrocarbons flash in the reboiler.

Careless handling when installing the packing can also cause powdering.

As particles break down, the pressure drop through the stripper increases.

This restricts the flow of vapor and liquid and causes the glycol to percolate out the top of the stripper.

Dirty packing, caused by sludge deposits of salt or tarry hydrocarbons, will also cause solution foaming in the stripper and increase glycol losses.

Packing should be cleaned or replaced when plugging or powdering occurs.

The same size tower packing should be used for replacement.

The standard size of the ceramic saddle or a stainless steel pall ring is one inch.

When stripping gas is used and a tower packing is placed in the downcomer between the reboiler and the storage tank, provisions should be made to replace the tower packing without cutting into the downcomer.

During low circulation rates:

Rich glycol may channel through the packing, causing poor contact between the liquid and hot vapors.

To prevent channeling, a distributor plate can be placed below the rich glycol feed line to evenly spread the liquid.

A large carryover of liquid hydrocarbons into the glycol system can be very troublesome and dangerous.

The hydrocarbons will flash in the reboiler, flood the stripper, and increase glycol losses.

Heavy hydrocarbon vapors and/or liquids could also spill over the reboiler and create a serious fire hazard.

Therefore, the vapors leaving the stripper vent should be piped away from the process equipment for safety.

The vent line should be properly sloped all the way from the stripper to the point of discharge to prevent condensed liquids from plugging the line.

If the vent line is long and is carried above the ground, a top vent, at a point not more than 20 ft away from the stripper, should probably be

installed to allow the escape of vapors in case of a freeze-up in the long line.

The piping should be the same size as the vessel connection, or larger. In areas where there is a possibility of cold, freezing weather, the vent line should be insulated from the stripper to the discharge point to prevent freeze-ups.

This will prevent the steam from condensing, freezing, and plugging the line.

If freezing occurs, the water vapor flashed in the reboiler may discharge into the storage tank and dilute the lean glycol.

The pressure caused by these trapped vapors could also force the regenerator to burst.

8.3.4.3 Reboiler

The reboiler supplies heat to separate the glycol and water by simple distillation.

Large plant locations may use hot oil or steam in the reboiler.

Remote field locations are generally equipped with a direct-fired heaters (Figure 8.25), with the following characteristics:

- Use a portion of the gas for fuel

- Heating element usually has a U-Tube shape and contains one or more burners

- Should be conservatively designed to:

 - Ensure long tube life

 - Prevent glycol decomposition caused by overheating (see Figures 8.26 and 8.27)

- Reboilers should be equipped with a high-temperature safety overriding controller to shut down the fuel supply gas system in case the primary temperature controller malfunctions

The firebox heat flux (a measure of the heat transfer rate in Btu/h/ft²) issues include the following:

- It should be high enough to provide adequate heating capacity but low enough to prevent glycol decomposition.

- Excessive heat flux, a result of too much heat in a small area, will thermally decompose the glycol (see Figure 8.28).

- The pilot flame should be kept low, especially in small reboilers for the following reasons:

 - To prevent glycol decomposition



Figure 8.25 Direct-fired reboiler.



Figure 8.26 Decomposed glycol on reboiler firetube.

To prevent tube burnout (see [Figure 8.29](#))

A low pilot flame is particularly important on smaller units where the pilot flame can supply a substantial portion of the total heat requirement

The pilot flame should be correctly adjusted to give a long, rolling, and slightly yellow-tipped flame

Nozzles are available that distribute the flame more evenly along the tube:

Decreases the heat flux of the area nearest the nozzle without actually lowering the total heat energy transferred



Figure 8.27 Decomposed glycol on reboiler firetube. *Courtesy of Gly-Tech.*

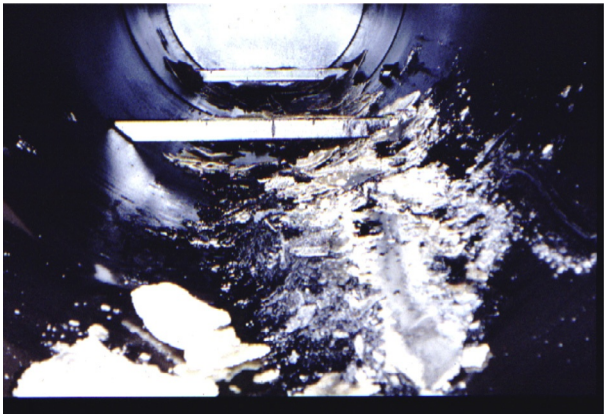


Figure 8.28 Decomposed glycol in reboiler vessel shell.



Figure 8.29 Decomposed glycol with firetube burnout.

Avoids direct and hard impingement of the flame against the fire tube

A pump shutdown device can prevent the circulation of wet glycol caused by a flame failure

A continuous spark ignition system, or a spark igniter to relight the pilot if it goes out, is also useful

Orifices on the air-gas mixers and pilots should be regularly cleaned to prevent burner failures

The following temperatures in the reboiler should not be exceeded to prevent burner failures:

Type of Glycol	Thermal Decomposition Temperatures
Ethylene	329 °F (165 °C)
Diethylene	328 °F (164 °C)
Triethylene	404 °F (207 °C)

Excessive discoloration and very slow degradation will result when the reboiler bulk temperature is maintained about 10 °F (5 °C) in excess of the above listed temperatures.

If coke, tarry products, and/or salt deposit on the fire tube, the heat transfer rate is reduced and a tube failure can result.

Localized overheating, especially where salt accumulates, will decompose the glycol.

An analysis of the glycol determines the amounts and types of these contaminants.

Salt deposits can also be detected by shutting off the burner on the reboiler at night and looking down the firebox.

A bright red-glowing light will be visible at spots on the tubes where salt deposits have collected.

These deposits can cause a rapid fire-tube burnout, particularly if the plant inlet separator is inadequate and a slug of salt water enters the absorber.

Coke and tarry products present in the circulating glycol can be removed by good filtration.

More elaborate equipment is needed to remove the salt.

Contaminates, which have already deposited on the fire tube and other equipment, can only be removed by using chemicals.

The heating process is thermostatically controlled and fully automatic.

The reboiler temperature should be occasionally verified with a test thermometer to make sure true readings are being recorded.

If the temperature fluctuates excessively when operating below the design capacity, the fuel gas pressure should be reduced.

A uniform temperature gives a better operation of the reboiler.

If the reboiler temperature cannot be raised as desired, it may be necessary to increase the fuel gas pressure up to about 30 psig.

If water and/or hydrocarbons enter the reboiler from the absorber, it may be impossible to raise the temperature until this problem is corrected.

Standard orifices furnished for reboiler burners are sized for 1000-1100 Btu/scf of gas.

If the rating of the fuel gas is less than this, it may be necessary to install a larger orifice or drill out the existing orifice to the next higher size.

Fires have been caused by leaks in the gas lines near the firebox.

The best precaution is to place valves and regulators in the gas line at a maximum distance from the firebox.

Another effective measure is the addition of a flame arrestor around the firebox.

If the arrestor is properly designed, even severe gas leaks in the immediate vicinity of the firebox will not ignite.

During a unit startup, it is imperative that the reboiler temperature be up to the desired operating level before flowing gas through the absorber.

The reboiler must be horizontal when erected.

A nonhorizontal position can cause a fire-tube burnout.

The reboiler should also be located close enough to the absorber to prevent excessive cooling of the lean glycol during cold weather. This will prevent hydrocarbon condensation and high glycol losses in the absorber.

8.3.4.4 Stripping Gas

Stripping gas is an optional item used to achieve very high glycol concentrations that cannot be obtained with normal regeneration.

It will provide the maximum dew-point depression and greater dehydration.

Stripping gas is used to remove the residual water after the glycol has been reconcentrated in the regeneration equipment.

It is also used to provide intimate contact between the hot gas and the lean glycol after most of the water has been removed by distillation.

Lean glycol concentrations in the range of 99.5–99.9% and dew-point depressions of 140 °F and above have been reported.

There are several methods of introducing stripping gas into the system.

One method is to use a vertical tray or packed section in the down-comer between the reboiler and the storage tank where the dry gas strips the additional water out of the regenerated glycol.

The glycol from the reboiler flows down through this section, contacts the stripping gas to remove the excess water, and goes into the storage tank.

Another method is to use glycol stripping gas sparger in the reboiler beneath the fire tube.

As the glycol flows through the reboiler, gas is injected into this vessel and is heated by the glycol.

Stripping gas contacts the glycol in the reboiler and removes some of the additional water.

Gas then passes out the stripper to the waste pit.

The lean glycol flows from the reboiler down into the storage tank. Stripper gas is normally taken from the reboiler fuel gas line (if dehydrated gas) at the fuel drip pot pressure.

Air or oxygen should not be used.

Stripping gas is usually controlled by a manual valve with a pressure gauge to indicate the flow rate through an orifice.

Stripping gas rate has the following characteristics:

It will vary according to the lean concentration desired and the method of glycol–gas contact

It is usually between 2 and 10 scf/gallon of glycol circulated

It should not get high enough to flood the stripper and blow glycol out to the pit

When stripping gas is used, it is necessary to provide for more reflux in the stripper to prevent excessive glycol losses.

This is usually provided by using a cool glycol condenser coil in the stripper.

8.3.4.5 Circulating Pump

A circulating pump is used to move glycol through the system.

The pump can be powered by electricity, gas, steam, or gas and glycol, depending upon the operating conditions and unit location.

A gas-glycol pump, which is a versatile piece of equipment, is commonly used for the following reasons:

The controls are serviceable, dependable, and, if adjusted properly, should provide long, trouble-free operation

It utilizes the rich glycol under pressure in the absorber to furnish part of its required driving energy

Since the pump cannot get more glycol back than it pumped over, a supplemental volume is needed to provide the driving force

Gas, under pressure from the absorber, is taken with the rich glycol to supply this additional volume

At 1000 psi operating pressure on the absorber, the volume of gas required is ~5.5 scf per gallon of lean glycol circulated

8.3.4.5.1 Helpful Maintenance Tips

Careful starting of a new pump can save much worry and downtime.

Pump packing gland generally used is lubricated only by the glycol itself.

Packing is dry when the pump is new.

As it soaks up glycol, the packing tends to expand.

If it has been screwed down too tight, either the packing will score the plunger or the packing will burn out.

The pump normally handles a fluid that is frequently dirty and corrosive.

Poor maintenance can lead to cylinder corrosion, seal erosion, impeller damage, pump cup or ring wear, and sticking or plugged valves.

These parts must be checked and kept in proper condition to keep the pump at maximum efficiency.

Pump rate should be commensurate with the gas volume being processed.

Speed should be decreased for low gas rates and increased for high rates.

Proportioning adjustments allow increased gas-glycol contact time in the absorber.

When the pump check valves become worn or clogged, the pump will operate normally except no fluid will go to the absorber.

Even a pressure gauge will indicate a pumping cycle.

The only evidence of this type of failure is little or no dew-point depression.

One sure way to check the volume flowing is to close the valve on the absorber outlet and calculate the flow by measuring the rise in the gauge glass (if one is available) versus the amount normally pumped

One of the most common sources of glycol loss occurs at the pump packing gland.

If the pump leaks more than one or two quarts of glycol per day, the packing needs to be replaced.

An adjustment will not recover the seal. Packing should be installed hand-tight and then backed off one complete turn.

If the packing gets too tight, the pistons can score and require replacement.

Glycol circulation rate of 2-3 gallons/lb of water to be removed is sufficient to provide adequate dehydration.

An excessive rate can overload the reboiler and reduce the dehydration efficiency.

The rate should be checked regularly by timing the pump to make sure it is running at the proper speed.

Proper pump maintenance will reduce the operating costs.

When the pump is not working, the whole system must be shut down because the gas cannot be dried effectively without a good continuous flow of glycol in the absorber.

Therefore, small replacement parts should be readily available to prevent lengthy shutdowns.

If there is insufficient glycol circulation:

Check the pump suction strainer for plugging and/or open the bleeder valve to eliminate air lock.

Glycol strainers should be regularly cleaned to avoid pump wear and other problems.

Pumps should be lubricated regularly.

Easy access to the pump can save time and trouble when making repairs and replacing components.

The maximum operating temperature of the pump is limited by the moving O-ring seals and nylon D slides.

A maximum temperature of 200 °F (94 °C) is recommended.

Packing life will be extended considerably if the temperature is held to a maximum of 150 °F (66 °C).

Therefore, sufficient heat exchange is necessary to keep the dry, lean glycol below these temperatures when it goes through the pump.

The pump is usually the most overworked and overused piece of equipment in the glycol process system.

The glycol system usually contains a second spare pump to avoid shutdowns when the primary pump fails.

It is not uncommon for operators to use the second pump to send more glycol to the absorber to avoid wet sales gas problems.

This procedure increases operating problems.

All of the other process variables should first be checked before a second pump is used.

A pressure gauge is furnished on the discharge side of the pump.

A valve is also furnished between the pressure gauge and the line so the pressure gauge can be isolated.

Pressure gauge can be used to see that the pump is working by watching the gauge “kick” as the pump piston strokes.

The sensing element in the pressure gauge is a bourdon tube.

The flexing, or movement, of this tube indicates the pressure.

A bourdon tube will fatigue or fail if subjected to continuous fluctuations in pressure on the pump discharge.

Pressure should be kept off the gauge except when testing the unit or to determine glycol loss from the gauge failure.

8.3.4.6 Flash Tank or Glycol-Gas Separator

The flash tank, or glycol-gas separator, is an optional piece of equipment used to recover the off-gas from the glycol-powered pump and the gaseous hydrocarbons from the rich glycol.

The recovered gas can be used as fuel to the reboiler and/or stripping gas.

Any excess gas is usually discharged through a back-pressure valve. The flash tank will keep volatile hydrocarbons out of the reboiler. This low-pressure separator may be located in one of the following two places:

Between the pump and the preheat coil in the storage tank

Between the preheat coil and the stripper

The separator usually works best in a temperature range of 130–170 °F (55–77 °C).

A two-phase separator, with at least a 5 min retention time, can be used to remove the gas.

If liquid hydrocarbons are present in the rich glycol, a three-phase separator should be used to remove these liquids before they get in the stripper and reboiler.

A liquid retention time of 20–45 min, depending on the type of hydrocarbons, API gravity, and the amount of foam, should be provided in the vessel.

Vessels should be located ahead of or behind the preheat coil in the storage tank, depending on the type of hydrocarbons present.

8.3.4.7 Gas Blanket

A gas blanket prevents air from contacting glycol in the reboiler and storage tanks.

A small amount of low-pressure gas is bled into the storage tank.

Gas is piped from the storage tank to the bottom of the stripper and it passes on overhead with the water vapor.

Elimination of air helps prevent glycol decomposition by slow oxidation.

The gas blanket equalizes the pressure between the reboiler and the storage tank. The gas blanket also prevents the liquid seal from breaking down between these two vessels.

8.3.4.8 Reclaimer

The reclaimer purifies the glycol for further use by vacuum distillation.

Clean glycol is driven off, and all the dirty sludge is left in the vessel and then washed to the sewer.

It is normally used only in very large glycol systems.



8.4 IMPROVING GLYCOL FILTRATION

8.4.1 General Considerations

Filters will do the following:

- Extend the life of pumps

- Prevent an accumulation of solids in the absorber (see [Figures 8.30–8.34](#))

- Prevent an accumulation of solids in the regeneration equipment

Solids that settle out on metal surfaces will frequently set up cell corrosion.

Filters remove the solids to also eliminate fouling, foaming, and plugging.

Filters should be designed to remove all solid particles 5 μm and larger. They should be able to operate up to a differential pressure of 20–25 psi without loss of seal or channeling of flow.

An internal relief valve with a setting of about 25 psi and differential pressure gauges are very helpful.

New elements should be installed before the relief valve opens.

For filters equipped with block and bypass valves:

- The bypass valve should be opened first before the block valves are closed to prevent excess pressure on the unit.

For filters not equipped with block and bypass valves:

- Close the block valve on the absorber glycol dump line before attempting to change elements.

Filters are not usually placed in the rich glycol line for best results, but the lean glycol can also be filtered to help keep the glycol clean.

- Frequent filter changes may be needed during unit startup or when neutralizers are added to control the glycol pH.

- New elements should be placed in a dry, clean place to keep them from dirt and grease.

Consult the filter manufacturer for installation and operating instructions.

- It is important to know when and how to change elements to keep air out the glycol system.

- Valves and gauges should be inspected occasionally for corrosion and scale buildup.

To determine the proper use of filter elements, cut them to the core and inspect them.

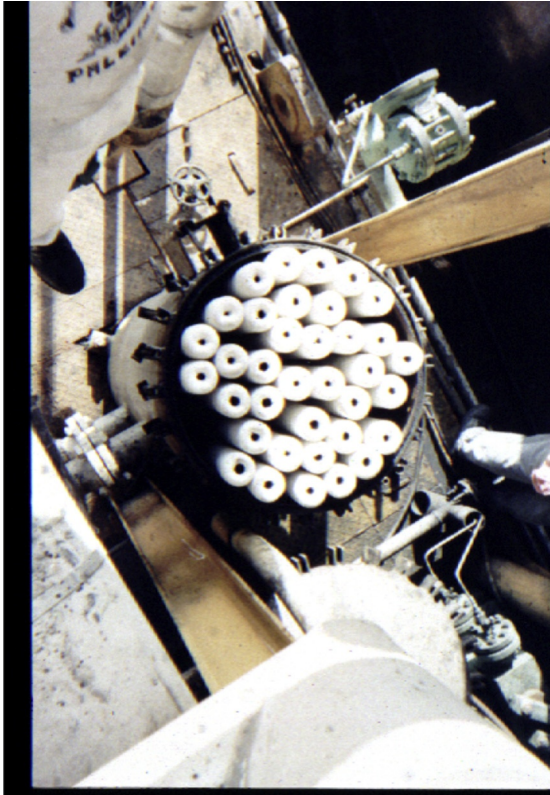


Figure 8.30 Top view of replacement of microfiber filter elements.

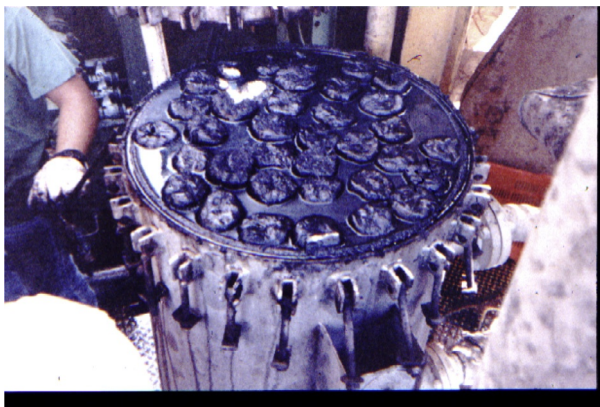


Figure 8.31 Microfiber filters contaminated with liquid hydrocarbon carryover.

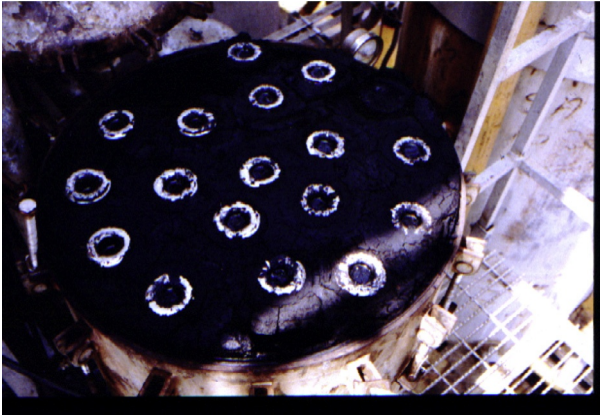


Figure 8.32 Decomposed glycol captured in the microfiber filters.



Figure 8.33 Decomposed glycol captured in the microfiber filters.



Figure 8.34 Filter collapsed due to high-pressure drop.

If they are dirty throughout, the filter is being used properly.
If the element is clean on the inside, an element with a different micron size may be needed.
It is also a good practice to occasionally scrape some sludge from a dirty element and have it analyzed.
This will help establish the types of contaminants present.
A record of the number of elements replaced will establish the amount of contaminants present.



8.5 USE OF CARBON PURIFICATION

8.5.1 General Considerations

Activated carbon can effectively eliminate most foaming problems by removing the hydrocarbons, well-treating chemicals, compressor oils, and other troublesome impurities from the glycol.

Two ways glycol purification can be achieved are as follows:

One method is to use two carbon towers installed in series but piped so they can be taken off-stream or interchanged without difficulty.
In large systems, about 2% of the total glycol flow should pass through the carbon towers.

In small systems, 100% of the total glycol flow should pass through the carbon towers.

Each carbon bed should be sized to handle 2 gallons of glycol per square foot of cross-sectional area per minute.

Towers should have an L/D ratio of about 3:1–5:1 and even 10:1 in some cases.

Towers should be designed to permit back-flushing with water to remove the dust after the carbon is loaded.

To achieve this, a retainer screen, with a smaller mesh size than the carbon should be installed above the carbon bed between the liquid inlet distributor and the outlet water drain nozzle to hold the carbon to the vessel.

The liquid distributor is needed to avoid glycol channeling through the carbon.

The screen size and support for the bottom of the towers should be carefully selected to avoid carbon plugging and to keep the carbon in the tower.

The inlet water nozzle for back-flushing should be placed below the screen in the bottom of the tower.

The appearance of the glycol can generally be used to determine when the carbon needs to be regenerated or replaced.

The pressure drop across the carbon bed can also be used.

The pressure drop normally across the carbon bed is only 1 or 2 lbs.

When the pressure drop reaches 10–15 lbs, the carbon is usually completely plugged with impurities.

Steam cleaning can sometimes be used to regenerate the carbon by removing the impurities.

However, this can be hazardous and offers only limited success.

Another method of purification is to use activated carbon in elements, such as Peco-Char.

Either purification system should be placed downstream from the solids filter.

This will increase the carbon adsorptive efficiency and life.



Gas Sweetening



9.1 PROCESSING NATURAL GAS

9.1.1 Introduction

Natural gas used by consumers is significantly different from the natural gas that is brought up from the wellhead.

The processing of natural gas, in many respects, is less complicated than the processing of crude oil but is equally important to the end users.

Natural gas used by consumers is composed almost entirely of methane.

Natural gas found at the wellhead, although composed primarily of methane, contains a number of impurities that need to be removed.

Raw natural gas comes from three types of wells:

- Oil wells
- Gas wells
- Condensate wells

Natural gas from oil wells is termed “associated gas.” The gas can exist separate from oil in formation and is called “free gas,” or it can be dissolved in the crude oil and is termed “dissolved gas.” Natural gas from gas and condensate wells contains little or no crude oil and is called “nonassociated gas.” Gas wells produce raw natural gas. Condensate wells produce free natural gas along with a semi-liquid hydrocarbon condensate.

Whatever the source of raw natural gas, it exists in mixtures with other associated hydrocarbons such as ethane, propane, butane, *i*-butane, and pentanes plus (natural gasoline). In addition to the associated hydrocarbons, raw natural gas also contains impurities such as water vapor, hydrogen sulfide (H_2S), carbon dioxide (CO_2), helium, nitrogen, and other compounds.

9.1.2 Natural Gas Processing

Processing consists of separating all of the various hydrocarbons and impurities from the raw natural gas to produce what is termed “pipeline quality” dry natural gas. Pipeline companies impose restrictions on the makeup of natural gas allowed into the pipeline. Pipeline companies require the raw

natural gas to be purified by removing “waste products” and associated hydrocarbons such as ethane, propane, butane, *i*-butane, and pentane plus (natural gasoline).

Associated hydrocarbons termed “natural gas liquids” (NGLs) can be a valuable byproduct of natural gas processing. They can be sold separately and have a variety of different uses, such as enhancing oil recovery in oil wells, providing raw materials for oil refineries and/or petrochemical plants, and as sources of energy.

Some of the needed processing can be accomplished at or near the wellhead (field processing). Complete processing of natural gas takes place at a processing plant, usually located in a natural gas producing region. Extracted natural gas is transported to processing plants through a network of gathering pipelines. A complex gathering system may consist of thousands of miles of pipeline, interconnecting the processing plant to more than 100 wells.

In addition to the processing done at the wellhead and centralized processing plants, some final processing is also done at “extraction plants.” These plants are located on major pipeline systems and remove small quantities of NGLs that may still exist in pipeline-quality gas.

Natural gas processing to meet pipeline-quality gas standards involves the following four main steps to remove the various impurities:

- Oil and condensate removal
- Water removal
- Separation of NGLs
- Sulfur and carbon dioxide removal

In addition to the four processes above, heaters and scrubbers are installed, usually at or near the wellhead. Scrubbers remove sand and other large-particle impurities. Heaters ensure that the temperature of the gas does not drop below the hydrate formation temperature.

When the gas temperature drops there is a tendency for “hydrates” to form. Hydrates are solid or semi-solid compounds, resembling ice-like crystals. Should these hydrates accumulate, they can impede passage of natural gas through valves and gathering systems, plug process instrumentation, and reduce the capacity of processing vessels. To reduce the occurrence of hydrates, the following equipment may be used:

- Indirect-fired heater
- Hydrate inhibitors
- Dehydration
- Low-temperature units

9.1.3 Oil and Condensate Removal

In order to process and transport associated dissolved natural gas, the gas must be separated from the oil in which it is dissolved. This is often accomplished by using equipment installed at or near the wellhead. Actual processes used to separate oil from natural gas and the equipment used can vary widely.

Dry pipeline-quality natural gas is basically identical across different geographic regions. Raw natural gas from different regions may have different compositions and separation requirements.

In some instances, natural gas is dissolved in oil underground primarily due to the pressure that the formation is under. When natural gas and oil is produced, it is possible that it will separate on its own, simply due to decreased pressure (similar to opening a bottle of soda and allowing the release of dissolved carbon dioxide). Separation of oil and gas is relatively easy, and the two hydrocarbons are separated and exit the separator for further processing. Conventional separators are used that provide gravity separation to separate the heavy liquids (oil and water) and lighter fluid (natural gas).

In some instances, specialized process equipment, such as a low-temperature separator (LTX), is necessary to separate oil and natural gas. The LTX is used for wells producing high-pressure gas along with light crude oil or condensate. These separators use pressure differentials to cool the wet natural gas and separate the oil and condensate. Wet gas enters the separator and is cooled slightly by a heat exchanger. The gas then travels through a high-pressure liquid “knockout,” which serves to remove any liquids into a LTX. Gas then flows into this LTX through a choke mechanism, which expands the gas as it enters the separator. The rapid expansion of the gas allows for the lowering of the temperature in the separator. After liquid removal, the dry gas then travels back through the heat exchanger and is warmed by the incoming wet gas. By varying the pressure of the gas in various sections of the separator, it is possible to vary the temperature, which causes the oil and some water to be condensed out of the wet gas stream. This basic pressure-temperature relationship can work in reverse as well to extract gas from a liquid oil stream.

9.1.4 Water Removal

In addition to separating oil and some condensate from the wet gas stream, it is necessary to remove most of the associated water. Most of the liquid, free water associated with extracted natural gas is removed by simple separation

methods at or near the wellhead. The removal of the water vapor that exists in solution in natural gas requires a more complex treatment. This treatment consists of “dehydrating” the natural gas, which usually involves one of the following two processes:

- Adsorption
- Absorption

Adsorption occurs when the water vapor is condensed and collected on the surface of a solid desiccant. Absorption occurs when the water vapor is taken out by a dehydrating agent, such as glycol.

9.1.5 Glycol Dehydration

An example of absorption dehydration is glycol dehydration. A liquid desiccant dehydrator serves to absorb water vapor from the gas stream. The principle agent in this process is glycol, which has a chemical affinity for water. When glycol comes in contact with a stream of natural gas that contains water, the glycol absorbs the water vapor out of the stream.

Glycol dehydration involves using a glycol solution, usually either diethylene glycol (DEG) or triethylene glycol (TEG), which is brought into contact with the wet gas stream in what is called a “contactor.” The glycol absorbs the water from the wet gas. Once absorbed, the glycol particles become heavier and sink to the bottom of the contactor where they are removed. The natural gas, having been stripped of most of its water content, is then transported out of the dehydrator.

The glycol solution, bearing all of the water stripped from the natural gas, is put through a specialized boiler designed to vaporize only the water out of the solution. While water has a boiling point of 212 °F (100 °C), glycol does not boil until 400 °F (204 °C). This boiling point differential makes it relatively easy to remove water from the glycol solution, allowing it to be reused in the dehydration process.

A new innovation in this process has been the addition of flash tank separator-condensers. As well as absorbing water from the wet gas stream, the glycol solution occasionally carries with it small amounts of methane and other compounds found in the wet gas. In the past, methane was simply vented out of the reboiler. In addition to losing a portion of the natural gas that was extracted, this venting contributes to air pollution and the greenhouse effect. In order to decrease the amount of methane and other compounds that are lost, flash tank separator-condensers work to remove these compounds before the glycol solution reaches the boiler.

A flash tank separator consists of a device that reduces the pressure of the glycol solution stream, allowing the methane and other hydrocarbons to vaporize or “flash.” The glycol solution then travels to the reboiler, which may also be fitted with air- or water-cooled condensers that serve to capture any remaining organic compounds that may remain in the glycol solution. In practice, these systems recover 90–99% of the methane that would otherwise be flared into the atmosphere.

9.1.6 Solid-Desiccant Dehydration

Solid-desiccant dehydration is the primary form of dehydrating natural gas using adsorption and usually consists of two or more adsorption towers, which are filled with a solid desiccant. Typical desiccants include activated alumina, silica gel, and molecular sieve.

Wet natural gas is passed through these towers from top to bottom. As the wet gas passes around the particles of desiccant material, water is retained on the surface of these desiccant particles. Passing through the entire desiccant bed, almost all of the water is adsorbed onto the desiccant material, leaving the dry gas to exit the bottom of the tower.

Solid-desiccant dehydrators are typically more effective than glycol dehydrators and are usually installed where very dry gas is required, such as upstream of a cryogenic expander, LPG, and LNG plants. These types of dehydration systems are best suited for large volumes of gas under very high pressure and are thus usually located on a pipeline downstream of a compressor station.

Two or more towers are required for this process due to the fact that after a certain period of use (typically 8 h), the desiccant in a particular tower becomes saturated with water. To “regenerate” the desiccant, a high-temperature heater is used to heat gas to a very high temperature. Passing this heated gas through a saturated desiccant bed vaporizes the water in the desiccant tower, leaving it dry and allowing for further natural gas dehydration.

9.1.7 Separation of NGLs

Natural gas coming directly from a well contains many NGLs that are commonly removed.

In most instances, NGLs have a higher value as separate products; thus, it is economical to remove them from the gas stream.

The removal of NGLs usually takes place in a relatively centralized processing plant and involves techniques similar to those used to dehydrate natural gas. There are two basic steps to the treatment of NGLs in the natural gas stream: first, the liquids must be extracted from the natural gas; and second, these NGLs must be separated themselves, down to their base components.

9.1.8 NGL Extraction

The two principle techniques for removing NGLs from the natural gas stream are the absorption method and the cryogenic expander process. These two processes account for nearly 90% of total NGL production.

9.1.8.1 Absorption Method

The absorption method of NGL extraction is very similar to using absorption for dehydration. The main difference is that, in NGL absorption, an absorbing oil is used as opposed to glycol. The absorbing oil has an “affinity” for NGLs in much the same manner as glycol has an affinity for water. Before the oil has picked up any NGLs, it is brought into contact with the absorption oil. As the natural gas is passed through an absorption tower, it is brought into contact with the absorption oil that soaks up a high proportion of the NGLs.

The “rich” absorption oil, now containing NGLs, exits the absorption tower through the bottom. It is now a mixture of absorption oil, propane, butanes, pentanes, and heavier hydrocarbons. The rich oil is fed into lean oil stills, where the mixture is heated to a temperature above the oil.

This process allows recovery of approximately

- 75% of butanes
- 85–90% pentanes and heavier molecules from the natural gas stream

This basic absorption process can be modified to improve its effectiveness or to target the extraction of specific NGLs. In the refrigerated oil absorption method, lean oil is cooled through refrigeration. Propane recovery can be upward of 90%, and 40% of ethane can be extracted from the natural gas stream. Extraction of the other heavier NGLs can be close to 100% using this process.

9.1.8.2 Cryogenic Expansion Process

Cryogenic processes are also used to extract NGLs from natural gas. Absorption methods can extract almost all of the heavier NGLs. Lighter hydrocarbons, such as ethane, are often more difficult to recover from the natural gas

stream. In certain instances, it is economically wise to simply leave the lighter NGLs in the natural gas stream. If it is economical to extract ethane and other lighter hydrocarbons, cryogenic processes are required for high recovery rates.

Cryogenic processes consist of dropping the temperature of the gas stream to around -120°F (-84°C). There are a number of different ways of chilling the gas to these temperatures. One of the most effective methods is the turbo-expander process. This process uses external refrigerants to cool the natural gas stream. An expansion turbine is used to rapidly expand the chilled gases, which causes the temperature to drop significantly. The rapid drop in temperature condenses ethane and other hydrocarbons in the gas stream and maintains methane in gaseous form.

This process allows the recovery of about 90–95% of the ethane. It converts some of the energy released when the natural gas stream is expanded into recompressing the gaseous methane effluent, thus saving energy costs associated with extracting ethane. The extraction of NGLs from the natural gas stream produces both cleaner, purer natural gas, as well as the valuable hydrocarbons that are the NGLs themselves.

9.1.9 NGL Fractionation

Once NGLs have been removed from the natural gas stream, they must be broken down into their base components to be useful. The mixed stream of different NGLs must be separated out. The process used to accomplish this task is called fractionation.

Fractionation is based on the different boiling points of the different hydrocarbons in the NGL stream. The process occurs in stages consisting of boiling off hydrocarbons one by one. The name of a particular fractionator gives an idea as to its purpose, as it is conventionally named for the hydrocarbon that is boiled off.

The entire fractionation process is broken down into steps, starting with the removal of the lighter NGLs from the stream. The particular fractionators are used in the following order:

- De-ethanizer—This step separates the ethane from the NGL stream.
- De-propanizer—This step separates the propane.
- De-butanizer—This step boils off the butanes, leaving the pentanes and heavier hydrocarbons in the NGLs.
- Butane splitter or de-isobutanizer—This step separates the *i*- and *n*-butanes.

By proceeding from the lightest hydrocarbons to the heaviest, it is possible to separate the different NGLs reasonably easily.

9.1.10 Sulfur and Carbon Dioxide Removal

In addition to water, oil, and NGL removal, one of the most important parts of gas processing involves the removal of sulfur and carbon dioxide. Natural gas from some wells may contain significant amounts of sulfur and carbon dioxide. Natural gas with hydrogen sulfide (H_2S) and other sulfur products is called “sour gas.” Sour gas is undesirable because sulfur compounds can be extremely harmful, even toxic if one breathes them, and they can also be extremely corrosive. Sulfur that exists in the gas stream can be extracted and marketed on its own. Fifteen percent of US sulfur production is obtained from gas processing plants.

9.1.11 Gas Sweetening Plant

Sulfur exists in natural gas as hydrogen sulfide (H_2S), and the gas is usually considered sour if the hydrogen sulfide content exceeds 5.7 mg of H_2S per cubic meter of natural gas. The process for removing hydrogen sulfide and carbon dioxide from a natural gas stream is referred to as “sweetening” the gas.

The primary process for sweetening natural gas is similar to processes of glycol dehydration and NGL absorption. Amine solutions are used to remove the hydrogen sulfide and carbon dioxide. The process is known simply as the “amine process,” and it is used in the majority of onshore gas sweetening operations. Gas with hydrogen sulfide and/or carbon dioxide is run through a tower, which contains the amine solution. This solution has an affinity for carbon dioxide and hydrogen sulfide and absorbs these contaminants much like glycol absorbing water.

The two principle amine solutions used are monoethanolamine (MEA) and diethanolamine (DEA). Either of these compounds, in liquid form, absorbs carbon dioxide and hydrogen sulfide from natural gas as it passes through. The effluent gas is virtually free of carbon dioxide and hydrogen sulfide compounds. Like the process for NGL extraction and glycol dehydration, the amine solution used can be regenerated (i.e., the absorbed sulfur is removed), allowing it to be reused to treat more gas.

Although many gas sweetening plants use the amine absorption process, it is also possible to use solid desiccants such as iron sponges and gas permeation. Extracted sulfur can be sold if it is reduced to its elemental form.

Elemental sulfur is a bright yellow powder-like material and can be seen in large piles near the treatment plants. In order to recover elemental sulfur from the gas processing plant, the sulfur-containing discharge from a gas sweetening process must be further treated.

The process used to recover sulfur is known as the Claus process and involves using thermal and catalytic reactions to extract the chemical sulfur from the hydrogen sulfide solution. The Claus process is usually able to recover 97% of the sulfur that has been removed from the natural gas stream. Once the natural gas has been fully processed and is ready to be consumed, it must be transported from those areas that produce natural gas to those areas that require it. The remainder of this chapter discusses in detail the processes used to sweeten natural gas streams containing carbon dioxide and hydrogen sulfide.



9.2 ACID GAS CONSIDERATIONS

Carbon dioxide (CO_2), hydrogen sulfide (H_2S), and other sulfur compounds, such as mercaptans, are known as acid gases and may require complete or partial removal to meet contract specifications.

9.2.1 Acid Gases

H_2S combined with water forms sulfuric acid. CO_2 combined with water forms carbonic acid.

Both are undesirable because they cause corrosion and reduce heating value and sales value. H_2S is poisonous and may be lethal. [Table 9.1](#) shows physiological effects of H_2S in air.

9.2.2 Sour Gas

Sour gas is defined as natural gas with H_2S and other sulfur compounds.

9.2.3 Sweet Gas

Sweet gas is defined as natural gas without H_2S and other sulfur compounds.

9.2.4 Gas Sales Contracts Limit Concentration of Acid Compounds

9.2.4.1 CO_2 Limitations

- 2–4% for pipelines
- Lowers Btu content

Table 9.1 Effects of H₂S Concentrations in Air Concentrations in Air

Percent by Volume	Parts per Million by Volume	Grains Per 100 Standard Cubic Feet ^a	Milligrams per Cubic Meter ^a	Physiological Effects
0.00013	0.13	0.008	0.18	Obvious and unpleasant odor is generally perceptible at 0.13 ppm and is quite noticeable at 4.6 ppm. As the concentration increases, the sense of smell fatigues, and the gas can no longer be detected by odor
0.002	10	1.26	28.83	Acceptable ceiling concentration permitted by federal OSHA standards
0.005	50	3.15	72.07	Acceptable maximum peak above the OSHA acceptable ceiling concentrations permitted once for 10 min per eight-hour shift, if no other measurable exposure occurs
0.01	100	6.30	144.14	Coughing, eye irritation, loss of sense of smell after 3 to 15 min. Altered respiration, pain in eyes, and drowsiness after 15–30 min, followed by throat irritation after one hour. Prolonged exposure results in a gradual increase in the severity of these symptoms
0.02	200	12.59	288.06	Rapidly kills sense of smell, burns eyes and throat
0.05	500	31.49	720.49	Dizziness, loss of sense of reasoning and balance. Breathing problems after a few minutes. Victims need prompt artificial resuscitation
0.07	700	44.08	1008.55	Unconsciousness. Breathing will stop and death will result if medical attention is not promptly received. Cardiorespiratory resuscitation is needed

Table 9.1 Effects of H₂S Concentrations in Air—cont'd
Concentrations in Air

Percent by Volume	Parts per Million by Volume	Grains Per 100 Standard Cubic Feet	Milligrams per Cubic Meter	Physiological Effects
0.10 +	1000 +	62.98	1440.98 +	Unconsciousness at once. Permanent brain damage or death may result without prompt medical attention and cardiorespiratory resuscitation

^aBased on 1% hydrogen sulfide = 629.77 gr/100 SCF at 14.696 psia and 59 F, or 101.325 kPa and 15C.

- CO₂ is corrosive
- 20 ppm for LNG plants

9.2.4.2 H₂S Limitations

- ¼ grain sulfur per 100 scf (~4 ppm)
- 0.0004% H₂S
- 2 ppm for LNG plants
- H₂S is toxic
- H₂S is corrosive (refer to NACE MR-01-75)

9.2.5 Partial Pressure

Partial pressure is used as an indicator if treatment is required. Partial pressure is defined as

$$PP = (\text{total pressure of system})(\text{mol\% of gas})$$

where CO₂ is present with water, a partial pressure >30 psia (207 kPa) would indicate CO₂ corrosion might be expected. Below 15 psia (103 kPa) would indicate CO₂ corrosion would not normally be a problem although inhibition may be required.

Factors that influence CO₂ corrosion are those directly related to solubility, that is, temperature, pressure, and composition of the water. Increased pressure increases solubility and increased temperature decreases solubility.

H₂S may cause sulfide stress cracking due to hydrogen embrittlement in certain metals. H₂S partial pressure >0.05 psia (0.34 kPa) necessitates treating.

9.2.6 NACE RP 0186

As shown in [Figure 9.1](#), National Association of Corrosion Engineers (NACE) Recommended Practice (RP) 0186 recommends special metallurgy to guard against H_2S .



9.3 SWEETENING PROCESSES

Numerous processes have been developed for acid gas removal and gas sweetening based on a variety of chemical and physical principles. [Table 9.2](#) lists the processes used to separate the acid gas from other natural gas components. The list, although not complete, represents many of the common available commercial processes ([Tables 9.3](#) and [9.4](#)).

Caution: Designers should consult with vendors and experts in acid gas treating before making a selection for any large plant.



9.4 SOLID BED PROCESSES

9.4.1 General Process Description

A fixed bed of solid particles can be used to remove acid gases either through chemical reactions or through ionic bonding.

This process flows the gas stream through a fixed bed of solid particles, which removes the acid gases and holds them in the bed.

When the bed is spent, the vessel must be removed from service and the bed regenerated or replaced. Since the bed must be removed from service to be regenerated, some spare capacity is normally provided.

Four commonly used processes under this category are:

- Iron sponge
- Sulfa-Treat
- Molecular sieve process
- Zinc oxide process

9.4.2 Iron Sponge Process

9.4.2.1 Application

The iron sponge process is economically applied to gases containing small amounts of H_2S (<300 ppm) operating at low to moderate pressures in the range of 50–500 psig (344.7–3447 kPa). This process does not remove CO_2 .

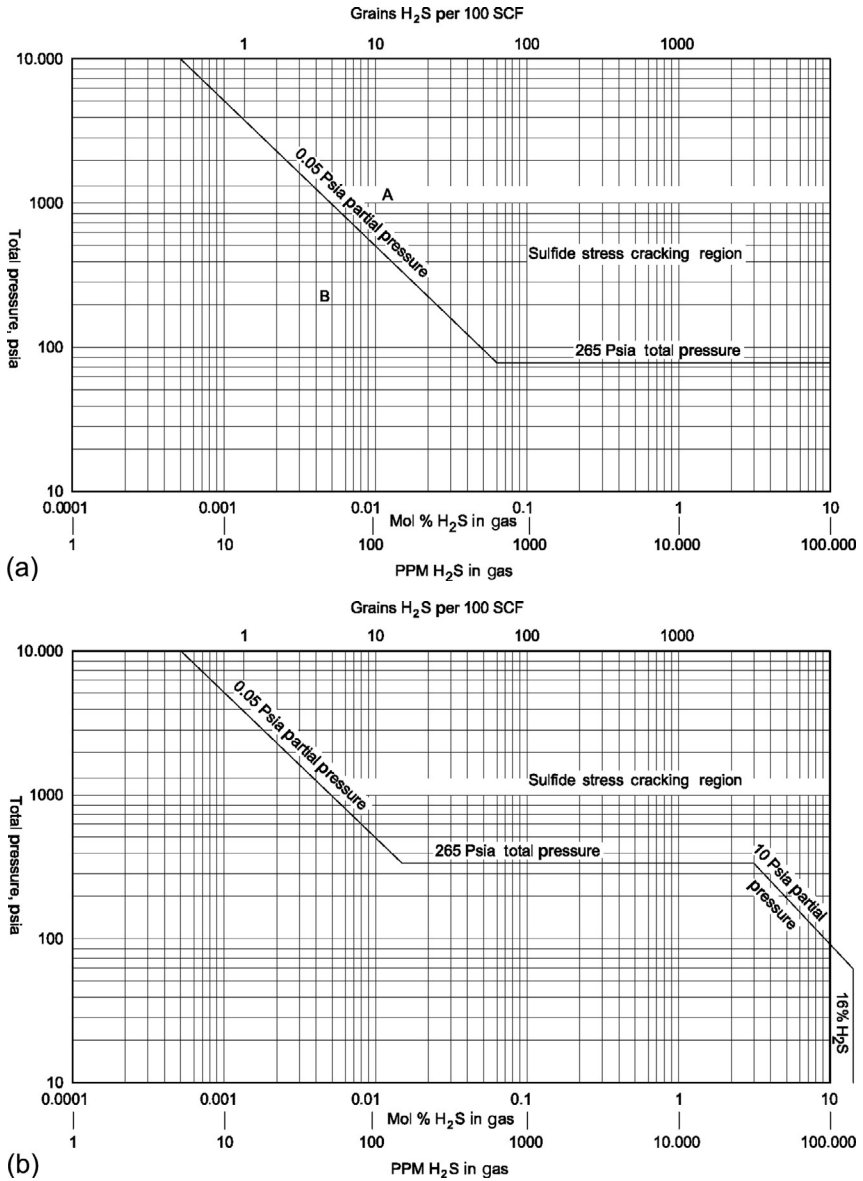
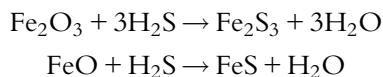


Figure 9.1 (a) Sulfide stress cracking regions in sour gas systems. (b) Sulfide stress cracking in multiphase systems.

Table 9.2 Acid Gas Removal Processes

Chemical Solvent	Physical Solvent	Direct Conversion
MEA	Selexol [®]	Iron Sponge
DEA	Rectisol	Stretford
TEA	Purisol	Unisulf
MDEA	Spasolv	Takahax
DIPA/Shell ADIP [®]	Propylene Carbonate	LO-CAT [®]
DGA/Fluor Econamine [®]	Estasolven	Lacy-Keller
Proprietary Amine	Alkazid	Townsend
Benfield (hot carbonate)		Sulfint
Catacarb (hot carbonate)		
Giammarco-Vetrocoke		
(hot carbonate)		
Diamox		
Dravo/Still		
Specialty Solvent	Distillation	Gas Permeation
Sulfinol [®]	Ryan Holmes	Membrane
Amisol	Cryofrac	Molecular Sieve
Flexsorb PS		
Selefining		
Ucarsol LE 711		
Optisol		
Zinc Oxide		
Sulfa-Check		
Slurrisweet		
Chemsweet		
Merox		

The reaction of iron oxide and H₂S produces iron sulfide and water as follows:



The reaction requires the presence of slightly alkaline water (pH 8-10) and a temperature below 110 °F (47 °C). When temperatures exceed 110 °F (47 °C), careful control of pH must be maintained. If the gas does not contain sufficient water vapor, water may need to be injected into the inlet gas stream. The pH level can be maintained through the injection of caustic soda, soda ash, lime, or ammonia with the water. pH control should be avoided whenever possible.

Table 9.3 Gases Removed by Various Processes

Process	Gases Removed				
	CO ₂	H ₂ S	RHS	COS	CS ₂
Solid Bed					
Iron Sponge		X			
Sulfa-Treat		X			
Zinc Oxide		X			
Molecular Sieves	X	X	X	X	X
Chemical Solvents					
MEA—MonoEthanolAmine	X	X		X ^a	X
DEA—DiEthanolAmine	X	X		X	X
MDEA—MethylDiEthanolAmine		X			
DGA—DiGlycolAmine	X	X		X	X
DIPA—DiIsoPropanolAmine	X	X		X	
Hot Potassium Carbonate	X	X		X	X
Physical Solvents					
Fluor Solvent	X	X	X	X	X
Shell Sulfinol [®]	X	X	X	X	X
Selexol [®]	X	X	X	X	X
Rectisol		X			
Direct Conversion of H ₂ S to Sulfur					
Claus		X			
LO-CAT [®]		X			
SulFerox [®]		X			
Stretford		X			
Sulfa-Check		X			
NASH		X			
Gas Permeation	X	X			

^aMEA reacts nonreversibly with COS (carbonyl sulfide), and, therefore, should not be used to treat gases with a large concentration of COS.

Although the presence of free alkalines enhances H₂S removal, it also creates potential safety hazards and promotes the formation of undesirable salts, adding to capital costs.

Ferric oxide is impregnated on wood chips, which produce a solid bed with a large ferric oxide surface area. Several grades of treated wood chips are available, based on iron oxide content. Ferric oxide wood chips are available in 6.5, 9.0, 15.0, and 20 lbs iron oxide/bushel. Chips are contained in a vessel, and sour gas flows downward through the bed and reacts with the ferric oxide.

Figure 9.2 shows a vertical vessel used in the iron sponge process.

Table 9.4 Process Capabilities for Gas Treating

	Normally Capable of Meeting 1/4 Grain ^a H ₂ S	Removes Mercaptans and COS Sulfur	Selective H ₂ S Removal	Solution Degraded (By)
Monoethanolamine	Yes	Partial	No	Yes (COS, CO ₂ , CS ₂)
Diethanolamine	Yes	Partial	No	Some (COS, CO ₂ , CS ₂)
Diglycolamine	Yes	Partial	No	Yes (COS, CO ₂ , CS ₂)
Methyldiethanolamine	Yes	Slight	Yes ^b	No
Sulfinol [®]	Yes	Yes	Yes ^b	Some (CO ₂ , CS ₂)
Selexol [®]	Yes	Slight	Yes ^b	No
Hot Pot-Benfield	Yes ^c	No ^d	No	No
Fluor Solvent	No ^e	No	No	No
Iron Sponge	Yes	Partial	Yes	-
Mol Sieve	Yes	Yes	Yes ^b	-
Stretford	Yes	No	Yes	CO ₂ at high concentrations
LO-CAT [®]	Yes	No	Yes	CO ₂ at high concentrations
Chemsweet	Yes	Partial for COS	Yes	No

^a1/4 grain H₂S/100 scf~ppmv H₂S.^bSome selectivity is exhibited by these processes.^cHi-pure version.^dHydrolizes COS only.^eCan be met with special design features.

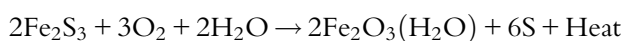
Source: GPSA Engineering Data Book, Tenth Edition, 1987.

9.4.2.2 Regeneration

Ferric sulfide can be oxidized with air to produce sulfur and regenerate the ferric oxide. Regeneration must be performed with care because the reaction with oxygen is exothermic (i.e., gives off heat).

Air must be introduced slowly so that the heat of reaction can be dissipated. If air is introduced quickly, the heat of reaction may ignite the bed. For this reason, spent wood chips should be kept moist when removed from the vessel. Otherwise, the reaction with oxygen in the air may ignite the chips and cause them to smolder.

The reactions for oxygen regeneration are as follows:



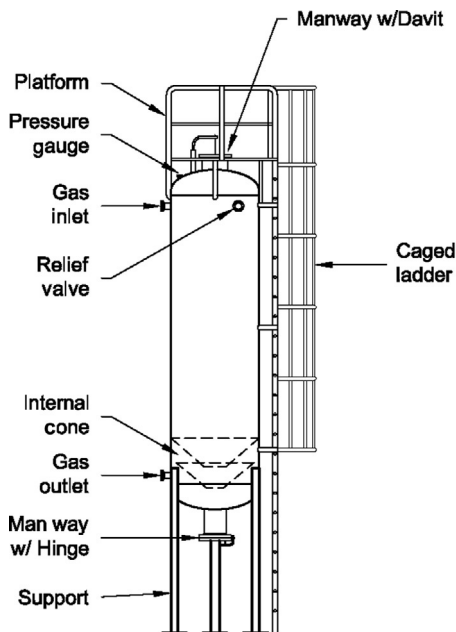
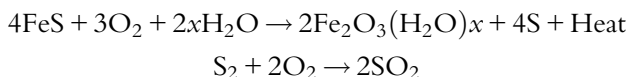


Figure 9.2 Iron oxide acid-gas treating unit.



Some of the elemental sulfur produced in the regeneration step remains in the bed. After several cycles, this sulfur will cake over the ferric oxide, decreasing the reactivity of the bed and causing excessive gas pressure drop. Typically after 10 cycles, the bed must be removed from the vessel and replaced with a new bed.

It is possible to operate an iron sponge with continuous regeneration by the introduction of small amounts of air in the sour gas feed. The oxygen in the air regenerates the iron sulfide and produces elemental sulfur. Although continuous regeneration decreases the amount of operating labor, it is not as effective as batch regeneration, and it may create an explosive mixture of air and natural gas. Due to the added costs associated with an air compressor, continuous regeneration generally does not prove to be the economic choice for the typically small quantities of gas involved.

9.4.2.3 Hydrate Considerations

Cooler operating temperatures of the natural gas, for example, during the winter, create the potential for hydrate formation in the iron sponge bed.

Hydrates can cause high-pressure drop, bed compaction, and flow channeling. When the potential for hydrates exists, methanol can be injected to inhibit their formation. If insufficient water is present to absorb the methanol, it may coat the bed, forming undesirable salts. Hydrocarbon liquids in the gas tend to accumulate on the iron sponge media, thus inhibiting the reactions.

The use of a gas scrubber upstream of the iron sponge and a gas temperature slightly less than that of the sponge media may prevent significant quantities of liquids from condensing and fouling the bed.

There has been a recent revival in the use of iron sponges to sweeten light hydrocarbon liquids. Sour liquids flow through the bed and are contacted with the iron sponge media and the reaction proceeds as above.

9.4.3 Sulfa-Treat Process

The Sulfa-Treat process similar to iron sponge process, utilizing the chemical reaction of ferric oxide with H_2S to sweeten gas streams.

9.4.3.1 Application

The Sulfa-Treat process is economically applied to gases containing small amounts of H_2S . Carbon dioxide is not removed in the process. This process utilizes a proprietary iron oxide co-product mixed with inert powder to form a porous bed. Sour gas flows through the bed and forms a bed primarily of pyrite. Powder has a bulk density of 70 lbs/ft^3 and ranges from 4 to 30 mesh.

The reaction works better with saturated gas and at elevated temperature up to 130°F (54.4°C). There is no minimum moisture or pH level required. The amount of bed volume required increases as the velocity increases and as the bed height decreases. Operation of the system below 40°F (4.4°C) is not recommended. Beds are not regenerated and must be replaced when the bed is spent.

9.4.4 Molecular Sieve Process

The molecular sieve process uses synthetically manufactured crystalline solids in a dry bed to remove gas impurities. Crystalline structure of the solids provides a very porous material having uniform pore size. Within the pores the crystalline structure creates a large number of localized polar charges called active sites. Polar gas molecules, such as H_2S and water vapor, which

enter the pores, form weak ionic bonds at the active sites. Nonpolar molecules, such as paraffin hydrocarbons, will not bond to the active sites.

Molecular sieves should be selected with a pore size that admits H_2S and water and prevents heavy hydrocarbons and aromatic compounds from entering. Carbon dioxide molecules are about the same size as H_2S molecules, but are nonpolar. CO_2 will enter the pores but will not bond to the active sites. Small quantities of CO_2 will be removed by becoming trapped in the pores by bonded H_2S or H_2O molecules blocking the pores. CO_2 will obstruct the access of H_2S and H_2O to the active sites, thus decreasing the overall effectiveness of the molecular sieve. Beds must be sized to remove all H_2O and provide for interference from other molecules in order to remove all H_2S .

The adsorption process usually occurs at moderate pressure. Ionic bonds tend to achieve an optimum performance near 450 psig (3100 kPa) but can operate in a wide range of pressures.

9.4.4.1 Regeneration

Regeneration is accomplished by flowing hot sweet stripping gas through the bed. Gas breaks the ionic bonds and removes the H_2S and H_2O . Typical regeneration temperatures are in the range of 300–400 °F (150–200 °C).

9.4.4.2 Mechanical Degradation

Care should be taken to minimize mechanical damage to the solid crystals as this will decrease the bed's effectiveness. The main cause of mechanical degradation is the sudden pressure and/or temperature changes that may occur when switching from adsorption to regeneration cycles. Proper instrumentation can significantly extend bed life.

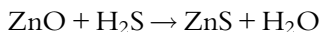
9.4.4.3 Application

The molecular sieve process is limited to small gas streams operating at moderate pressures. It is generally used for polishing applications following one of the other processes.

9.4.5 Zinc Oxide Process

9.4.5.1 Process

The equipment used in the zinc oxide process is similar to the iron sponge process. It uses a solid bed of granular zinc oxide to react with the H_2S to form zinc sulfide and water as shown below



The rate of reaction is controlled by the diffusion process, as the sulfide ion must first diffuse to the surface of the zinc oxide to react. Temperatures above 250 °F (120 °C) increase the diffusion rate, which promotes the reaction rate. The strong dependence on diffusion means that other variables, such as pressure and gas velocity, have little effect on the reaction.

9.4.5.2 Bed Considerations

The zinc oxide is contained in long thin beds to lessen the chances of channeling. The pressure drop through the beds is low. Bed life is a function of gas H₂S content and can vary from 6 months to more than 10 years. Beds are often used in series to increase the level of saturation prior to change out of the catalyst. A spent bed is discharged by gravity flow through the bottom of the vessel.

9.4.5.3 Application

The zinc oxide process is seldom used due to disposal problems with the spent bed, which is classified as a heavy metal salt.



9.5 CHEMICAL SOLVENT PROCESSES

9.5.1 General Process Description

Chemical solvent processes utilize an aqueous solution of a weak base to chemically react with and absorb the acid gases in the natural gas stream. Absorption occurs as result of the partial pressure differential between the gas and the liquid phases.

9.5.1.1 Regeneration

Reactions are reversible by changing the system temperature, pressure, or both. Aqueous base solution can be regenerated and circulated in a continuous cycle.

9.5.1.2 Most Common Chemical Solvents

The most common chemical solvents are:

- Amines
- Carbonates

9.5.2 Amine Processes

9.5.2.1 Amine Considerations

Several processes have been developed using the basic action of various amines. Amines are categorized by the number of organic groups bonded to the central nitrogen atom, as primary, secondary, or tertiary. Primary amines form stronger bases than secondary amines, which form stronger bases than tertiary amines. Amines with stronger base properties are more reactive toward CO_2 and H_2S gases and form stronger chemical bonds. This implies that the acid gas vapor pressure will be lower for a given loading as amine reactivity increases and a higher equilibrium loading may be achieved.

9.5.2.2 Process Description

A typical amine system is shown in Figure 9.3. Sour gas enters the system through an inlet scrubber to remove any entrained water or hydrocarbon liquids. Gas enters the bottom of the amine absorber and flows counter-current to the amine solution. The absorber tower consists of trays (diameters >20 in. (500 mm)), conventional packing (diameters <20 in. (500 mm)) or structured packing (diameters >20 in. (500 mm)). Sweetened gas leaves the top of the tower. An optional outlet scrubber may be included to recover entrained amine from the sweet gas. Since the natural gas leaving

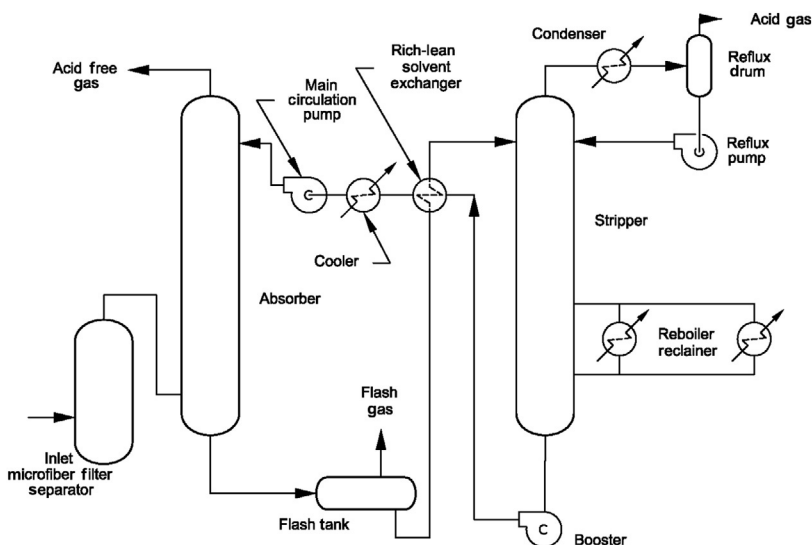


Figure 9.3 Gas sweetening process flow schematic (amine sweetening).

the top of the tower is saturated with water, the gas will require dehydration before entering a pipeline.

Rich amine, solution containing CO_2 and H_2S , leaves the bottom of the absorber and flows to the flash tank where most of the dissolved hydrocarbon gases or entrained hydrocarbon condensates are removed. A small amount of the acid gases flash to the vapor phase. From the flash tank, the rich amine proceeds to the rich amine/lean amine heat exchanger where it recovers some of the sensible heat from the lean amine stream, which decreases the heat duty on the amine reboiler and the solvent cooler. Preheated rich amine then enters the amine stripping tower where heat from the reboiler breaks the bonds between the amine and acid gases. Acid gases are removed overhead and lean amine is removed from the bottom of the stripper.

Hot lean amine flows to the rich amine/lean amine heat exchanger and then to additional coolers, typically aerial coolers, to lower its temperature to $\sim 10^\circ\text{F}$ (5.5°C) above the inlet gas temperature. This reduces the amount of hydrocarbons condensed in the amine solution when the amine contacts the sour gas.

A side stream of amine, of $\sim 3\%$, is taken off after the rich/lean amine heat exchanger and is flowed through a charcoal filter to clean the solution of contaminants. Cooled lean amine is then pumped up to the absorber pressure and enters the top of the absorber. Amine solution flows down the absorber where it absorbs the acid gases. Rich amine is then removed at the bottom of the tower and the cycle is repeated.

Most common amine processes are:

- Monoethanolamine
- Diethanolamine

Both processes remove CO_2 and H_2S to pipeline specifications.

9.5.2.3 Methyldiethanolamine

Methyldiethanolamine (MDEA) is a newer process used for selective removal of H_2S in the presence of CO_2 . It significantly reduces treating costs when CO_2 reduction is not necessary.

9.5.2.4 Monoethanolamine Systems

9.5.2.4.1 General Discussion

MEA is a primary amine, which has had widespread use as a gas sweetening agent. The process is well proven and can meet pipeline specifications. MEA is a stable compound and, in the absence of other chemicals, suffers no degradation or decomposition at temperatures up to its normal boiling point.

9.5.2.4.2 Regeneration

MEA reactions are reversible by changing the system temperature. Reactions with CO_2 and H_2S are reversed in the stripping column by heating the rich MEA to $\sim 245^\circ\text{F}$ at 10 psig (118°C at 69 kPa). Acid gases evolve into the vapor and are removed from the still overhead. Thus, the MEA is regenerated.

9.5.2.4.3 Disadvantages

MEA reacts with carbonyl sulfide (COS) and carbon disulfide (CS_2) to form heat-stable salts, which cannot be regenerated at normal stripping column temperatures. At temperatures above 245°F (118°C) a side reaction with CO_2 exists that produces oxazolidone-2, a heat-stable salt, which consumes MEA from the process. Normal regeneration temperature in the still will not regenerate heat-stable salts or oxazolidone-2. A reclaimer is often included to remove these contaminants.

9.5.2.4.4 Reclaimer

A side stream of from 1% to 3% of the MEA circulation is drawn from the bottom of the stripping column. The stream is then heated to boil the water and MEA overhead while the heat-stable salts and oxazolidone-2 are retained in the reclaimer. The reclaimer is periodically shut in and the collected contaminants are cleaned out. When the contaminants are removed from the system, any MEA bonded to them is also lost.

9.5.2.4.5 Solution Concentration and Solution Loading

Both the solution concentration and solution loading are limited to avoid excessive corrosion. MEA is usually circulated in a solution of 15–20% MEA by weight in water. Operating experience indicates that the solution loading should not be $>0.3\text{--}0.4$ mol of acid gas per mole of MEA and is largely determined by the $\text{H}_2\text{S}/\text{CO}_2$ ratio. The greater the ratio (i.e., the higher the concentration of H_2S relative to CO_2), the higher the allowable loading and amine concentration. This is due to the reaction of H_2S and iron (Fe) to form iron sulfide (Fe_2S_3 and FeS), which forms a protective barrier on the steel surface. This barrier can be stripped away by high fluid velocities and may lead to increased corrosion on the exposed steel.

9.5.2.4.6 Corrosion Considerations

Acid gases in the rich amine are corrosive, but the above concentration limits may hold corrosion to acceptable levels. Corrosion commonly shows up in

areas of carbon steel that have been stressed, such as heat affected zones near welds, areas of high acid gas concentration, or at a hot gas and liquid interface. Thus, stress relieving all equipment after manufacturing is necessary to reduce corrosion, and special metallurgy is usually used in specific areas such as the still overhead or the reboiler tubes.

9.5.2.4.7 Foam Considerations

MEA systems foam rather easily resulting in excessive amine carryover from the absorber. Foaming can be caused by a number of foreign materials such as condensed hydrocarbons, degradation products, solids such as carbon or iron sulfide, excess corrosion inhibitor, and valve grease, to name a few.

9.5.2.4.8 Microfiber Filter Separator

A microfiber filter separator should be installed at the gas inlet to the MEA contactor. It is an effective method of foam control and removes many of the contaminants before they enter the system. Hydrocarbon liquids are usually removed in the flash tank. Degradation products are removed in a reclaimer as described above.

9.5.2.4.9 Blanket Gas System

A gas blanket system is installed on MEA storage tanks and surge vessels. This prevents oxidation of MEA. Sweet natural gas or nitrogen is normally used.

9.5.2.4.10 MEA Losses

MEA has the lowest boiling point and the highest vapor pressure of the amines. This results in MEA losses of 1–3 lbs/MMSCF (16–48 kg/MM m³) of inlet gas.

9.5.2.4.11 Summary

MEA systems can effectively treat sour gas to pipeline specifications. Care in the design and material selection of MEA systems is required to minimize equipment corrosion.

9.5.2.5 DEA Systems

9.5.2.5.1 General Discussion

DEA is a secondary amine and is also used to treat natural gas to pipeline specifications. As a secondary amine, DEA is less alkaline than MEA. DEA systems do suffer the same corrosion problems, but not as severely as those using MEA. Solution strengths are typically from 25% to 35% DEA

by weight in water. DEA has significant advantages over MEA when COS or CS₂ are present. DEA reacts with COS and CS₂ to form compounds that can be regenerated in the stripping column. Thus, COS and CS₂ are removed without a loss of DEA.

9.5.2.5.2 Reclaimer

High CO₂ levels have been observed to cause DEA degradation to oxazolidones. DEA systems usually include a carbon filter but do not typically include a reclaimer due to the small amount of degradation product.

9.5.2.5.3 Solution Concentration and Solution Loading

Stoichiometry reactions of DEA and MEA with CO and H₂S are the same. Molecular weight of DEA is 105 compared to 61 for MEA. Combination of molecular weights and reaction stoichiometry means that ~1.7 lbs (0.77 kg) of DEA must be circulated to react with the same amount of acid gas as 1.0 lbs (0.45 kg) of MEA. The solution strength of DEA ranges up to 35% by weight compared to 20% for MEA. Loadings for DEA systems range from 0.35 to 0.65 mol of acid gas per mole of DEA without excessive corrosion. The result of this is that the circulation rate of a DEA solution is slightly less than in a comparable MEA system.

9.5.2.5.4 Amine Losses

Vapor pressure of DEA is ~1/30 of the vapor pressure of MEA. Thus, DEA amine losses are much lower than in an MEA system.

9.5.2.6 *Di-glycol-amine Systems*

9.5.2.6.1 General Discussion

DGA is a primary amine used in the Fluor Econamine process to sweeten natural gas. Reactions of DGA with acid gases are the same as those for MEA. Unlike MEA, degradation products from reactions with COS and CS₂ can be regenerated.

9.5.2.6.2 Solution Concentration and Solution Loading

DGA systems typically circulate a solution of 50–70% DGA by weight in water. At the above solution strengths and a loading of up to 0.3 mol of acid gas per mole of DGA, corrosion in DGA systems is slightly less than in MEA systems.

9.5.2.6.3 Advantages

Low vapor pressure decreases amine losses. High solution strength permits lower circulation rates.

9.5.2.7 Diisopropanolamine Systems

9.5.2.7.1 General Discussion

Diisopropanolamine (DIPA) is a secondary amine used in the Shell “**ADIP**” process to sweeten natural gas. It is similar to DEA systems but offers the following advantages: carbonyl sulfide (COS) can be removed and the DIPA solution can be regenerated easily, and the system is generally noncorrosive and has a lower energy consumption.

9.5.2.7.2 Advantages

At low pressures, DIPA will preferentially remove H_2S . As pressure increases, the selectivity of the process decreases and DIPA removes increasing amounts of CO_2 . Thus, this system can be used either to selectively remove H_2S or to remove both CO_2 and H_2S .

9.5.2.8 MDEA Systems

9.5.2.8.1 General Discussion

MDEA is a tertiary amine, which, like the other amines, is used to sweeten natural gas streams. Major advantage over other amine processes is MDEA can be used to selectively remove H_2S in the presence of CO_2 . If the gas is contacted at pressures ranging from 800 to 1000 psig (5500–6900 kPa), H_2S levels can be reduced to concentrations required by pipelines. While at the same time, 40–60% of the CO_2 present flows through the contactor, untreated.

9.5.2.8.2 $\text{CO}_2/\text{H}_2\text{S}$ Ratio

In cases where a high $\text{CO}_2/\text{H}_2\text{S}$ ratio is present, MDEA can be used to improve the quality of the acid gas stream to a Claus recovery plant, but the higher CO_2 content of the treated residue gas must be tolerated.

9.5.2.8.3 Solution Concentration and Solution Loading

Solution strengths typically range from 40% to 50% MDEA by weight. Acid gas loading varies from 0.2 to 0.4 or more moles of acid gas per mole of MDEA depending on the supplier. MDEA has a molecular weight of 119. MDEA solution makeup is dependent upon the supplier. It can be adjusted to optimize treatment for a particular gas inlet composition.

9.5.2.8.4 Advantages

Higher allowable MDEA concentration and acid gas loading results in reduced circulation flow rates. Significant capital savings are realized due to reduced pump and regeneration requirements. MDEA has a lower heat requirement due to its low heat of regeneration. In some applications, energy requirements for gas treating can be reduced as much as 75% by changing from DEA to MDEA.

9.5.2.9 *Inhibited Amine Systems*

9.5.2.9.1 General Discussion

Inhibited amine processes use standard amines that have been combined with special inhibiting agents that minimize corrosion. They allow higher solution concentrations and higher acid gas loadings, thus reducing required circulation rates and energy requirements. They also utilize hot potassium carbonate to remove CO_2 and H_2S . As a general rule, this process should be considered when the partial pressure of the acid gas is 20 psia (138 kPa) or greater. The process is not recommended for low-pressure absorption, or high-pressure absorption of low-concentration acid gas.

9.5.3 Hot Potassium Carbonate Systems

9.5.3.1 *General Discussion*

Potassium bicarbonate (KHCO_3) solutions are not readily regenerable in the absence of CO_2 .

Thus, these processes are only employed for H_2S removal when quantities of CO_2 are present.

Potassium carbonate also reacts reversibly with COS and CS_2 .

9.5.3.2 *Process Description*

Figure 9.4 shows a typical hot carbonate system for gas treating. Gas to be treated enters the bottom of the absorber and flows counter-current to the potassium carbonate. Sweet gas exits the top of the absorber. The absorber is typically operated at 230 °F (110 °C). Gas/gas exchanger may be included to cool the sweet gas, recover sensible heat, and decrease the system's utility heat requirements.

Rich potassium carbonate solution from the bottom of the absorber flows to a flash drum where much of the acid gas is removed. The solution then proceeds to the stripping column, which operates at ~245 °F (118 °C) and near atmospheric pressure. Low pressure, combined with a small amount of heat input, strips the remaining acid gases. Lean potassium

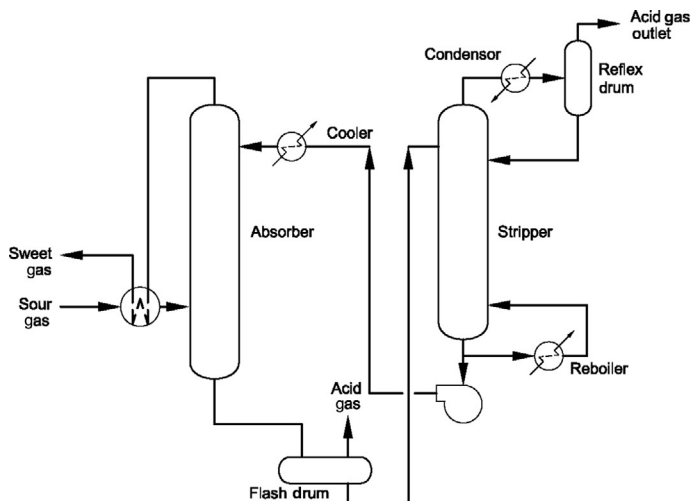


Figure 9.4 Gas sweetening flow schematic of a hot carbonate process.

carbonate from the stripper is pumped back to the absorber. Lean solution may or may not be cooled slightly before entering the absorber. Heat of reaction from the absorption of the acid gases causes a slight temperature rise in the absorber.

Solution concentration is limited by both solubility of potassium carbonate in the lean stream and solubility of the potassium bicarbonate (KHCO_3) in the rich stream. Reaction with CO_2 produces two moles of KHCO_3 per mole of potassium carbonate reacted. Thus, the KHCO_3 in the rich stream normally limits the lean solution potassium carbonate concentration to 20–35% by weight.

9.5.3.3 Performance

Potassium carbonate works best on gas streams with a CO_2 partial pressure of 30–90 psi (207–620 kPa). When CO_2 is not present, H_2S removal will be limited because the regeneration of the potassium carbonate requires an excess of KHCO_3 . The presence of CO_2 in the gas provides a surplus of KHCO_3 in the rich stream. Pipeline-quality gas often requires secondary treating using an amine or similar system to reduce H_2S level to 4 ppm.

9.5.3.4 Dead Spot Considerations

Because this system is operated at high temperatures to increase the solubility of carbonates, the designer must be careful to avoid dead spots in the system

where the solution could cool and precipitate solids. If solids do precipitate, the system may suffer from plugging, erosion, or foaming.

9.5.3.5 Corrosion Considerations

Hot potassium carbonate solutions are corrosive. All carbon steel must be stress-relieved to limit corrosion. A variety of corrosion inhibitors, such as fatty amines or potassium dichromate, are available to decrease corrosion rates.

9.5.4 Proprietary Carbonate Systems

Several proprietary processes add a catalyst, or activator. The catalyst increases the performance of the hot potassium carbonate system and increases the reaction rates both in the absorber and in the stripper. In general, these processes also decrease corrosion in the system. Some of the proprietary processes for hot potassium carbonate include

Benfield: Several activators

Girdler: Alkanolamine activators

Catacarb: Alkanolamine and/or borate activators

Giammarco-Vetrocoke: Arsenic and other activator

9.5.5 Specialty Batch Chemical Solvents

9.5.5.1 General Discussion

Several batch chemical processes have been developed and have specific areas of application.

Processes include

- Zinc oxide slurry
- Caustic wash
- Sulfa-Check
- Slurrisweet
- Chemsweet

9.5.5.1.1 Process Description

Gas is flowed into a vessel and contacted with the solvent. Acid components are converted to soluble salts, which are nonregenerable, limiting the life of the solution.

Once saturation levels are reached, the solution must be replaced.

For some of these processes, the spent solutions are not hazardous, but for others, the spent solutions have been labeled hazardous and, if used, must be disposed of as Class IV materials.

9.5.5.2 Performance

Units in these processes have a wide operating range, with acid gas concentrations ranging from as low as 10 ppm to as high as 20%. Operating pressures range from near atmospheric to >1000 psig (7000 kPa). Some units have been designed to handle from several thousand cubic feet per day to more than 15 MMSCFD (several hundred cubic meters per day to more than 420,000 m³ per day).

9.5.5.3 Sulfa-Check

Sulfa-Check is a single-step process that converts H₂S to sulfur in a bubble tower filled with a proprietary solution of oxidizing and buffering agents. Oxidizing agent is a proprietary formulation of chelated nitrite ions.

9.5.5.4 Concentration Considerations

Reaction rate is independent of the concentration of the oxidizing agent. There is no limit to the concentration of H₂S treated. Process is most economical for acid gas streams containing from 1 ppm to 1% H₂S. pH must be held above 7.5 to control selectivity and optimize H₂S removal. One gallon (4 l) of oxidizing solution can remove up to 2 lbs (1 kg) of H₂S when the system is operated at ambient temperatures <100 °F (38 °C). If gas temperatures exceed 100 °F (38 °C), the solubility of sulfur in the oxidizing agent decreases.

9.5.5.5 Bubble Flow

Operating pressure of at least 20 psig (138 kPa) is required for proper unit operation to maintain bubble flow through the column. Bubble flow is necessary to produce intimate mixing of the gas and liquid.

9.5.5.6 Disposal of Oxidizing Solution

Oxidizing solution will eventually become saturated and require replacement. Disposal of this slurry poses no environmental problem, as the reaction produces an aqueous slurry of sulfur and sodium salt.



9.6 PHYSICAL SOLVENT PROCESSES

9.6.1 General Process Description

Physical solvent systems are similar to chemical solvent systems but are based on the gas solubility within a solvent instead of a chemical reaction. Acid gas solubility depends on acid gas partial pressure and system temperature.

Higher acid gas partial pressures increase the acid gas solubility. Low temperatures increase acid gas solubility, but, in general, temperature is not as critical as pressure.

Various organic solvents are used to absorb the acid gases based on partial pressures. Regeneration of the solvent is accomplished by flashing to lower pressures and/or stripping with solvent vapor or inert gas. Some solvents can be regenerated by flashing only and require no heat. Other solvents require stripping and some heat, but typically the heat requirements are small compared to chemical solvents.

Physical solvent processes have a high affinity for heavy hydrocarbons. If the natural gas stream is rich in C^{3+} hydrocarbons, then the use of a physical solvent process may result in a significant loss of the heavier mole weight hydrocarbons. These hydrocarbons are lost because they are released from the solvent with the acid gases and cannot be economically recovered.

Physical solvent processes should be considered for gas sweetening under the following circumstances:

- Partial pressure of the acid gases in the feed is 50 psi (345 kPa) or higher
- Concentration of heavy hydrocarbons in the feed is low
- Only bulk removal of acid gases is required
- Selective H_2S removal is required

Figure 9.5 shows a physical solvent process.

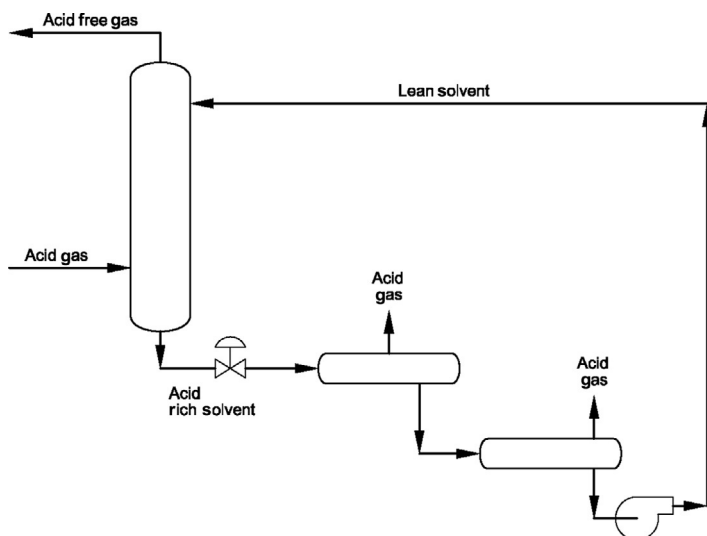


Figure 9.5 Typical flow schematic for a physical solvent process.

Reviewing the flow in [Figure 9.5](#), sour gas contacts the solvent using counter-current flow in the absorber. Rich solvent from the absorber bottom is flashed in stages to near atmospheric pressure. This causes the acid gas partial pressures to decrease, and the acid gases evolve to the vapor phase and are removed. Regenerated solvent is then pumped back to the absorber.

[Figure 9.5](#) is an example where flashing is sufficient to regenerate the solvent. Some solvents require a stripping column just prior to the circulation pump.

Some systems require temperatures below ambient, thus refrigeration using power turbines replaces the pressure reducing valves. These turbines recover some of the power from the high pressure rich solvent and thus decrease the utility power requirements for refrigeration and circulation. The majority of the physical solvent processes are proprietary and are licensed by the company that developed them.

Four typical processes are discussed below.

9.6.2 Fluor Solvent Process

The Fluor Solvent process uses propylene carbonate as a physical solvent to remove CO_2 and H_2S . Propylene carbonate also removes C^{3+} hydrocarbons, COS , SO_2 , CS_2 , and H_2O from the natural gas stream. Thus, in one step the natural gas can be sweetened and dehydrated to pipeline quality. This process is used for bulk removal of CO_2 and is not used to treat to $<3\% \text{CO}_2$.

This system requires special design features such as larger absorbers and higher circulation rates to obtain pipeline quality and usually is not economically applicable for outlet requirements.

Propylene carbonate has the following characteristics, which make it suitable as a solvent for acid gas treating:

- High degree of solubility for CO_2 and other gases
- Low heat of solution for CO_2
- Low vapor pressure at operating temperature
- Low solubility for light hydrocarbons (C_1 , C_2)
- Chemically nonreactive toward all natural gas components
- Low viscosity
- Noncorrosive toward common metals

The above characteristics combine to yield a system that has low heat and pumping requirements, is relatively noncorrosive, and suffers only minimal solvent losses ($<1 \text{ lbs/MMSCF}$).

Solvent temperatures below ambient are often used to increase solvent gas capacity and, therefore, decrease circulation rates. Expansion of the rich solvent and flash gases through power turbines can provide the required refrigeration. Alternately, auxiliary refrigeration may be included to further decrease circulation rates.

9.6.3 Sulfinol[®] Process

The Sulfinol process was developed and licensed by Shell. It employs both a chemical and a physical solvent for the removal of H_2S , CO_2 , mercaptans, and COS. Sulfinol[®] solution is a mixture of tetrahydrothiophene dioxide (Sulfolane[®]), a physical solvent, and DIPA, a secondary amine, and water. DIPA, previously discussed, is the chemical solvent. Solution concentrations range between 25% and 40% sulfolane, 40–55% DIPA, and 20–30% water and depend on the conditions and composition of the gas being treated.

9.6.3.1 Acid Gas Loadings

The presence of the physical solvent, SulfolaneW, allows higher acid gas loadings compared to systems based on amine only. Typical loadings are 1.5 mol of acid gas per mole of Sulfinol solution. Higher acid gas loadings, together with a lower energy of regeneration, can result in lower capital and energy costs per unit of acid gas removed as compared to the ethanolamine processes.

9.6.3.2 Features

Features of the Sulfinol process include essentially complete removal of mercaptans, high removal rate of COS, lower foaming tendency, lower corrosion rates, and the ability to slip up to 50% CO_2 .

9.6.3.3 Design Considerations

The design is similar to that of the ethanolamines. Degradation of DIPA to oxazolidones (DIPA-OX) usually necessitates the installation of a reclaimer for their removal.

9.6.3.4 Foam Considerations

As with the ethanolamine processes, aromatics and heavy hydrocarbons in the feed gas should be removed prior to contact with the Sulfinol solution to minimize foaming.

9.6.3.5 Factors to Consider Before Selecting a Treating Process

The merits of the Sulfinol process, as compared to the ethanolamine processes, are many, but there are other factors that must be considered before selecting the appropriate gas treating process. For example, a licensing fee, while not necessary for the ethanolamine processes, is required for the Sulfinol process. The solvent costs are generally higher for the Sulfinol process than they are for DEA. Operators are more familiar with DEA and the typical problems associated with this process. In cases of low acid gas partial pressure, the advantage of a lower circulation rate for the Sulfinol process diminishes compared to the DEA process.

9.6.4 Selexol[®] Process

The Selexol process uses the dimethylether of polyethylene glycol as a solvent. It is licensed by UOP. The process is selective toward removing sulfur compounds. Levels of CO₂ can be reduced by ~85%. The process is economical when high acid gas and partial pressures exist and there is an absence of heavy ends in the gas.

This process will not typically remove enough CO₂ to meet pipeline gas requirements. DIPA can be added to the solution to remove CO₂ down to pipeline specifications. The process also removes water to <7 lbs/MMSCF (0.11 g/std m³). The system then functions much like the Sulfinol process discussed earlier. The addition of DIPA increases the relatively low stripper heat duty.

9.6.5 Rectisol Process

The Rectisol process was developed by the German Lurgi Company and Linde A. G. It uses methanol to sweeten natural gas. Due to the high vapor pressure of methanol, this process is usually operated at temperatures of 30 to -100 °F (-34 to -74 °C). It has been applied for the purification of gas for LNG plants and in coal gasification plants, but is not commonly used to treat natural gas streams.



9.7 DIRECT CONVERSION PROCESSES

9.7.1 General Process Description

Chemical and physical solvent processes remove acid gas from the natural gas stream but release H₂S and CO₂ when the solvent is regenerated. The release of H₂S to the atmosphere is limited by environmental regulations. Acid gases

could be routed to an incinerator/flare, which would convert the H_2S to SO_2 . Environmental regulations restrict the amount of SO_2 vented or flared.

Direct conversion processes use chemical reactions to oxidize H_2S and produce elemental sulfur. These processes are generally based either on the reaction of H_2S and O_2 or H_2S and SO_2 . Both reactions yield water and elemental sulfur. These processes are licensed and involve specialized catalysts and/or solvents.

9.7.2 Stretford Process

9.7.2.1 General Discussion

The Stretford process uses O_2 to oxidize H_2S . It was originally licensed by the British Gas Corporation and is no longer in use.

9.7.2.2 Process Description

Figure 9.6 shows a simplified diagram of the Stretford process. The gas stream is washed with an aqueous solution of sodium carbonate, sodium vanadate, and anthraquinone disulfonic. An oxidized solution is delivered from the pumping tank to the top of the absorber tower where it contacts the gas stream in a counter-current flow.

The bottom of the absorber tower consists of a reaction tank from which the reduced solution passes to the solution flash drum, which is situated above the oxidizer. The reduced solution passes from here into the base of the oxidizer vessel. Hydrocarbon gases, which have been dissolved in the solution at the plant pressure, are released from the top of the flash drum.

Air is blown into the oxidizer, and the main body of the solution, now reoxidized, passes into the pumping tank. The sulfur is carried to the top of

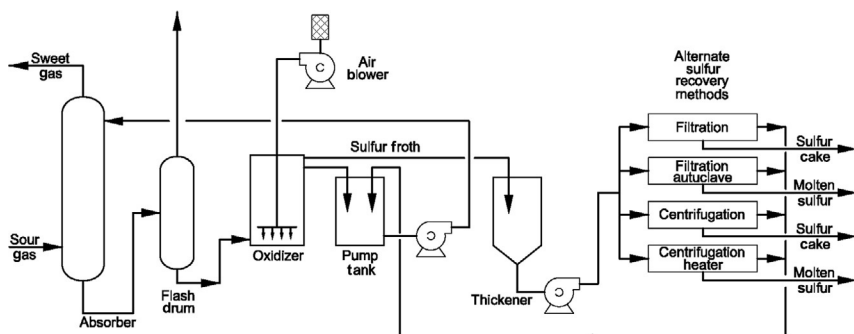
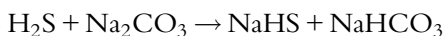


Figure 9.6 Simplified flow schematic of the Stretford process.

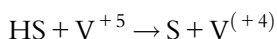
the oxidizer by froth created by the aeration of the solution and passes into the thickener.

The function of the thickener is to increase the weight percent of sulfur that is pumped to one of the alternate sulfur recovery methods of filtration, filtration and autoclaves, centrifugation or centrifugation with heating.

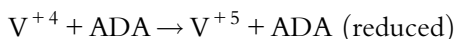
Chemical reactions involved are:



Sodium carbonate provides the alkaline solution for initial adsorption of H_2S and the formation of hydrosulfide (HS). The hydrosulfide is reduced in a reaction with sodium meta vanadate to precipitate sulfur



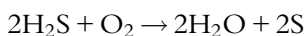
Anthraquinone disulfonic acid (ADA) reacts with 4-valent vanadium and converts it back to 5-valent



Oxygen from the air converts the reduced ADA back to the oxidized state as shown below:



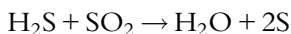
The overall reaction is



9.7.3 IFP Process

9.7.3.1 General Discussion

The IFP process was developed by the Institut Francais du Petrole. The process reacts H_2S with SO_2 to produce water and sulfur. The overall reaction is



9.7.3.2 Process Description

Figure 9.7 shows a simplified diagram of the IFP process. The process involves mixing the H_2S and SO_2 gases and then contacting them with a liquid catalyst in a packed tower. Elemental sulfur is recovered in the bottom of the tower. A portion of this must be burned to produce the SO_2 required to remove the H_2S .

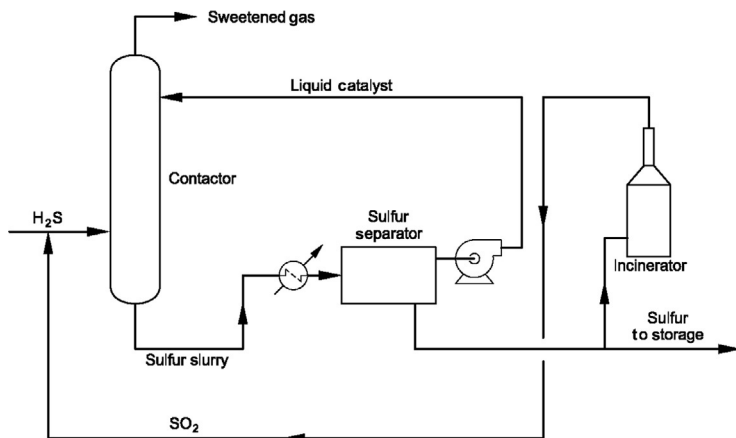


Figure 9.7 Simplified flow schematic of the IFP process.

9.7.3.3 Ratio of H_2S to SO_2

The most important variable is the ratio of H_2S to SO_2 in the feed. The ratio is controlled by analyzer equipment to maintain the system performance.

9.7.4 LO-CAT[®]

9.7.4.1 General Discussion

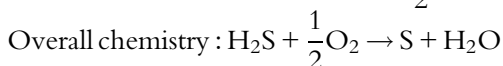
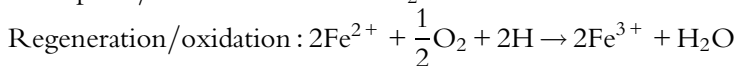
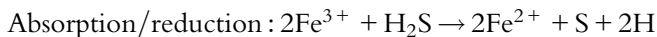
The LO-CAT process was developed by ARI Technologies (LO-CAT) and Shell. The process (Sulferox) combined development under the Merichem Company marketed as LO-CAT.

Processes employ high iron concentration reduction-oxidation technology for the selective removal of H_2S (not reactive to CO_2) to <4 ppm in both low and high pressure gas streams.

9.7.4.2 Process Description

Acid gas stream is contacted with the solution where H_2S reacts with and reduces the chelated iron and produces elemental sulfur. The iron is then regenerated by bubbling air through the solution. Heat is not required for regeneration.

The reactions involved are exothermic (give off heat):



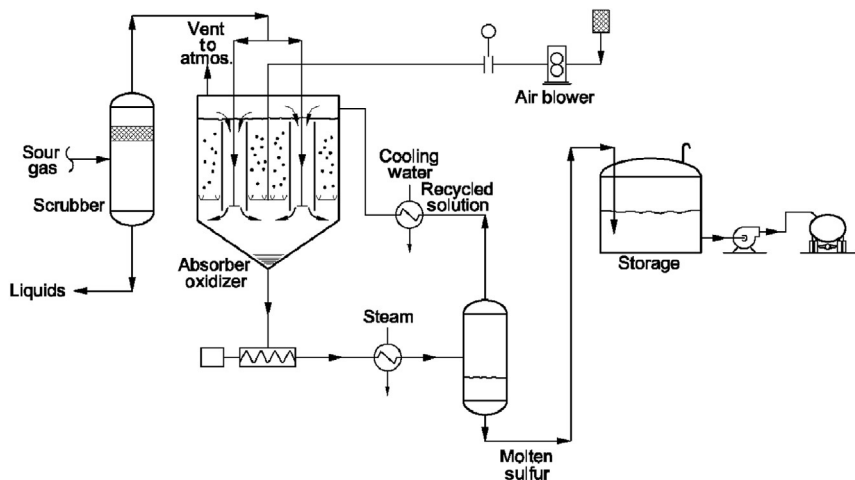


Figure 9.8 Simplified flow schematic of the LO-CAT process.

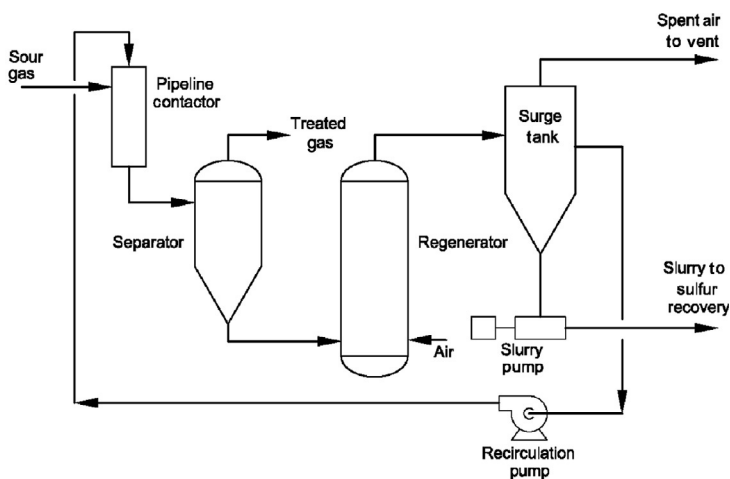


Figure 9.9 Simplified flow schematic of the sulferox process.

Figures 9.8 and 9.9 show a flow schematic for the LO-CAT[®] process.

9.7.4.3 Operating Considerations

The turndown is 100%. Solution is nontoxic; thus, no special disposal problems exist. No sulfur products are dispersed to the atmosphere. The process requires a centrifuge and slurry handling equipment.

9.7.4.4 Sulferox[®]

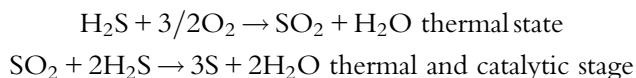
The Sulferox system uses a patented pipeline contactor with co-current flow to minimize sulfur plugging.

9.7.5 Claus

9.7.5.1 General Discussion

The Claus process is used to treat gas streams containing high (above 50%) concentrations of H_2S . Chemistry of the units involves partial oxidation of hydrogen sulfide to sulfur dioxide, and the catalytically promoted reaction of H_2S and SO_2 to produce elemental sulfur.

Reactions are staged and are as follows:



9.7.5.2 Process Description

Figure 9.10 shows a simplified flow diagram of a two-stage Claus process plant. The first stage of the process converts H_2S to sulfur dioxide and to sulfur by burning the acid gas stream with air in the reaction furnace. This provides SO_2 for the next phase of the reaction.

Gases leaving the furnace are cooled to separate out elemental sulfur formed in the thermal stage. Reheating, catalytically reacting, and sulfur

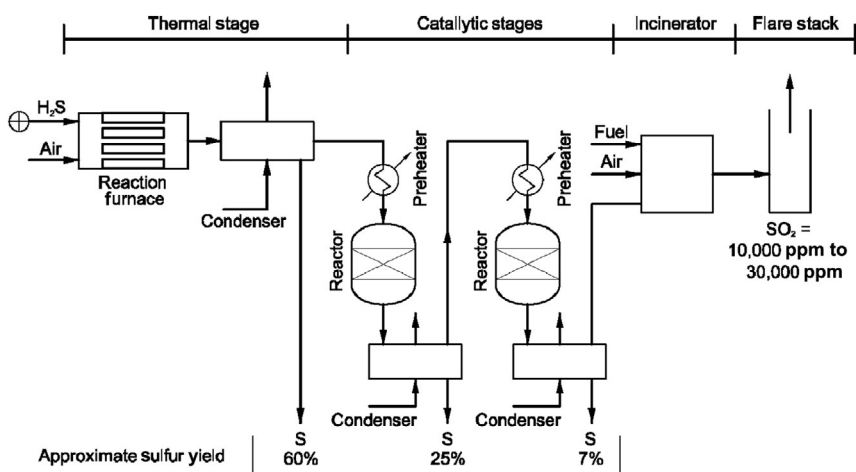


Figure 9.10 Simplified process flow schematic for a two-stage Claus process plant.

condensation removes additional sulfur. Multiple reactors are provided to achieve a more complete conversion of the H_2S . Condensers are provided after each reactor to condense the sulfur vapor and separate it from the main stream.

Conversion efficiencies of 94–95% can be attained with two catalytic stages while up to 97% conversion can be attained with three catalytic stages. The efficiencies are dictated by environmental concerns; the effluent gas (SO_2) is either vented, incinerated, or sent to a “tail gas treating unit.”

9.7.6 Tail Gas Treating

9.7.6.1 General Discussion

Many different processes are used in tail gas treating. Processes can be grouped into two categories. Leading among these processes are the Sulfreen and the Cold Bed Absorption processes. These processes are similar. The first utilizes two parallel Claus reactors in a cycle, where one reactor operates below the sulfur dew point to absorb the sulfur, while the second is regenerated with heat to recover the absorbed sulfur. Recoveries up to 99.9% of the inlet sulfur stream are possible.

The second category involves the conversion of the sulfur compounds to H_2S and then absorbing the hydrogen from the stream. The SCOT process appears to be the leading choice among this type of process. It uses an amine to remove the H_2S , which is usually recycled back to the Claus plant.

Other types of processes oxidize the sulfur compounds to SO_2 and then convert the SO_2 to a secondary product such as ammonium thiosulfate, a fertilizer. These plants can remove more than 99.5% of the sulfur and may eliminate the need for incineration. Costs of achieving tail gas cleanup are high—typically double the cost of a Claus unit.

9.7.7 Sulfa-Check

9.7.7.1 General Discussion

Sulfa-Check converts H_2S to sulfur in a bubble tower of oxidizing and buffering agents. It is used in applications with low H_2S concentrations and where regeneration is not required. The pH must be held above 7.5 and requires temperatures below 110°F (42°C), pressures above 20 psig (1.35 barg) and 2 lbs of H_2S per gallon of solution at temperatures $<110^\circ\text{F}$ (42°C).



9.8 DISTILLATION PROCESS

9.8.1 Ryan-Holmes Distillation Process

9.8.1.1 General Discussion

Distillation processes use cryogenic distillation to remove acid gases from a gas stream. The process is applied to remove CO₂ for LPG separation or where it is desired to produce CO₂ at high pressure for reservoir injection or other use.

9.8.1.2 Process Description

The process consists of two, three, or four fractionating columns. The gas stream is first dehydrated and then cooled with refrigeration and/or pressure reduction.

9.8.1.2.1 Three-Column System

The three-column system is used for gas streams containing <50% CO₂. The first column operates at 450–650 psig (3100–4500 kPa) and separates a high-quality methane product in the overhead. Temperatures in the overhead are from 0 to –140 °F (–18 to –95 °C). The second column operates at a slightly lower pressure and produces a CO₂ stream overhead, which contains small amounts of H₂S and methane. The bottom product contains H₂S and the ethane plus components. The third column produces NGL liquids, which are recycled back to the first two columns. It is this recycle that allows the process to be successful. NGL liquids prevent CO₂ solid formation in the first column and aid in the breaking of the ethane/CO₂ azeotrope in the second column to permit high ethane recoveries.

9.8.1.2.2 Four-Column System

The four-column system is used where CO₂ feed concentration exceeds 50%. The initial column in this scheme is a de-ethanizer. The overhead product, a CO₂/methane binary, is sent to a bulk CO₂ removal column and de-methanizer combination. CO₂ is produced as a liquid and is pumped to injection or sales pressure.

9.8.1.2.3 Two-Column System

The two-column system is used when a methane product is not required and is thus produced with the CO₂. Very high propane recoveries may be achieved; however, little ethane recovery is achieved. These processes

require feed gas preparation in the form of compression and dehydration, which adds to their cost. Such systems are finding applications in enhanced oil recovery (EOR) projects.



9.9 GAS PERMEATION PROCESS

9.9.1 Membranes

9.9.1.1 Definition

Membranes are thin semipermeable barriers that selectively separate some compounds from others.

9.9.1.2 Applications

Membranes are used primarily for bulk CO₂ removal from natural gas streams. Natural gas upgrading can remove CO₂ and H₂O to pipeline specifications.

9.9.1.2.1 Lowers H₂S levels

Membranes are used onshore/offshore, at wellhead, or at gathering facilities.

9.9.1.2.2 EOR Operations

Membranes are used to recover CO₂ from EOR floods for recycle injection into oil reservoir, thus increasing oil recovery CO₂ is removed from an associated natural gas stream and is reinjected into an oil to enhance oil recovery.

9.9.1.2.3 Materials Used for CO₂ Removal

Materials used for CO₂ removal include:

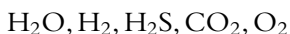
- Polymer based (properties modified to enhance performance)
- Cellulose acetate (most rugged)
- Polyimides, polyamides, polysulfone
- Polycarbonates, polyetherimide

9.9.1.3 Membrane Permeation

Membranes do not act as filters where small molecules are separated from larger ones through a medium of pores. They operate on the principle of solution-diffusion through a nonporous membrane. Highly solubilized components dissolve and diffuse through the membrane.

Relative permeation rates

- Most soluble (fastest gases)



- Least soluble (slowest gases)



CO_2 first dissolves into the membrane and then diffuses through it. Membranes allow selective removal of fast gases from slow gases.

9.9.1.3.1 Membranes Do Not Have Pores

Membranes do not separate on the basis of molecular size. Separation is based on how well different compounds dissolve into the membrane and then diffuse it.

Fick's law (known as Basic Flux Equation) is used to approximate the solution-diffusion process. It is expressed as

$$J = (k \times D \times \Delta p) / I \quad (9.1)$$

where J = membrane flux of CO_2 (rate/unit area); k = solubility of CO_2 in the membrane; D = diffusion coefficient of CO_2 through the membrane; Δp = partial pressure difference of CO_2 between the feed (high pressure) and permeate (low pressure) side of the membrane; I = membrane thickness.

The solubility (k) and diffusion (D) coefficients are usually combined into a new variable called permeability (P).

Fick's law can be split into two portions:

- Membrane dependent portion (P/I)
- Process dependent portion (Δp)

High flux requires:

- Correct membrane material
- Correct processing conditions

P/I is not constant and is sensitive to

- Pressure
- Temperature

Flick's Law equation can be equally written for methane or any other component in the stream. This leads to the definition of a second important variable called selectivity (σ).

9.9.1.3.2 Selectivity (σ)

The selectivity is the ratio of the permeabilities of CO_2 to other components in the stream. It is a measure of how much better the membrane permeates CO_2 to the compound in question. Example: a CO_2 -to-methane selectivity of 30 means that CO_2 permeates the membrane 30 times faster than methane.

9.9.1.3.3 Important Parameters When Selecting a Membrane

Permeability (P)

High permeability results in less membrane area required for a given separation and a lower system cost.

Selectivity (α)

High selectivity results in lower losses of hydrocarbons as CO_2 is removed and a higher volume of salable product.

Unfortunately, high CO_2 permeability does not correspond to high selectivity. A choice must be made between a highly elective, or permeable, membrane and somewhere between on both parameters. Unfortunately, high CO_2 permeability does not correspond to high selectivity. The usual choice is to use a highly selective material and then make it as thin as possible to increase the permeability. Reduced thickness makes the membrane extremely fragile and therefore unusable.

In the past, membrane systems were not a viable process because the membrane thickness required to provide the mechanical strength was so high that the permeability was minimal.

9.9.1.4 Asymmetric Membrane Structure

An asymmetric membrane structure features a **single polymer** consisting of an extremely thin **nonporous layer** mounted on a much thicker and **highly porous layer** of the same material, as opposed to a homogenous structure, where membrane porosity is more-or-less uniform throughout.

Figure 9.11 is an example of an asymmetric membrane.

Nonporous layer

Meets the requirements of the ideal membrane, that is, highly selective, and thin.

Porous layer

Provides mechanical support and allows the free flow of compounds that permeate through the nonporous layer.

9.9.1.5 Composite Membrane Structure

The disadvantages of the asymmetric membrane structure are they are composed of a single polymer; they are expensive to make out of exotic, highly customized polymers; and they are produced in small quantities. These drawbacks are overcome by producing a composite membrane. The composite membrane consists of a thin selective layer made of one polymer mounted on an asymmetric membrane, which is made of another polymer.

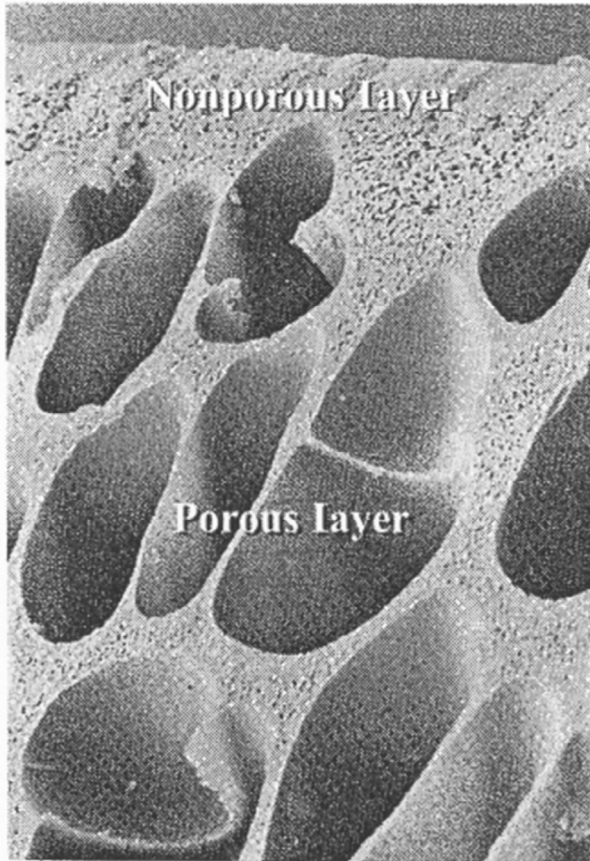


Figure 9.11 Asymmetric membrane structure.

The composite structure allows manufacturers to use readily available materials for the asymmetric portion of the membrane and specially developed polymers, which are highly optimized for the required separation and the selective layer. [Figure 9.12](#) is an example of a composite membrane structure.

Composite structures are being used in most newer advanced CO₂ removal membranes because the properties of the selective layer can be adjusted readily without significantly increasing membrane cost.

9.9.1.6 Membrane Elements

Membrane elements are manufactured in one of two forms:

- Flat sheet
- Hollow fiber

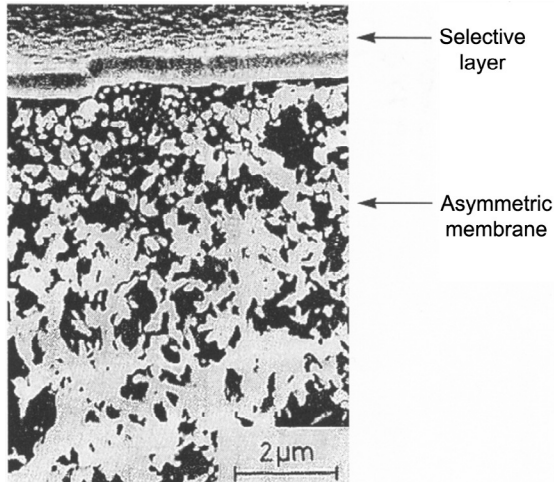


Figure 9.12 Composite membrane structure.

9.9.1.6.1 Flat Sheet

As shown in [Figure 9.13](#), the flat sheet is combined into a spiral-wound element. Two flat sheets of membrane with a permeate spacer in between are glued along three of their sides to form an envelope (or leaf) that is open at one end. Envelopes are separated by feed spacers and wrapped around a permeate tube with their open ends facing the permeate tube. Feed enters along the side of the membrane and passes through the feed spacers separating the envelopes. As the gas travels between the envelopes, CO_2 , H_2S , and other highly permeable compounds permeate into the envelope. Permeated components have only one outlet: they must travel within the envelope to the permeate tube. The driving force for transport is the low permeate and high-feed pressures.

9.9.1.6.2 Permeate gas

Permeate gas enters the permeate tube through holes drilled in the tube. The permeate gas travels down the tube to join the permeate from other tubes. Any gas on the feed side that does not get a chance to permeate, leaves through the side of the element opposite the feed position.

9.9.1.6.3 Optimization Considerations

Optimization involves the number of envelopes and element diameter.

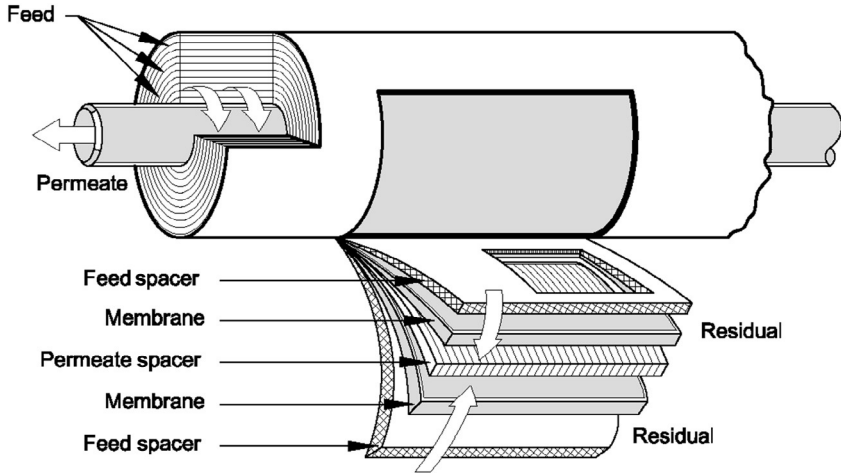


Figure 9.13 Spiral wound membrane element.

Number of envelopes

The permeate gas must travel the length of each envelope. Having many shorter envelopes makes more sense than having a few longer ones because pressure drop is greatly reduced in the former case.

Element diameter

A larger bundle diameter allows better packing densities but increases the element tube size and decreases cost. A larger diameter also increases the element weight, which makes the elements more difficult to handle during installation and replacement.

9.9.1.6.4 Hollow Fiber

As shown in [Figure 9.14](#), very fine hollow fibers are wrapped around a central tube in a highly dense pattern. In this wrapping pattern, both open ends of the fiber end up at a permeate pot on one side of the element. Feed gas flows over and between the fibers and some components permeate into them.

Permeate gas

Permeate gas travels within the fibers until it reaches the permeate pot, where it mixes with the permeate from other fibers. The total permeate exits the element through a permeate pipe. Gas that does not permeate eventually reaches the element's center tube, which is perforated in a way that is similar to that of the spiral-wound permeate tube. In this

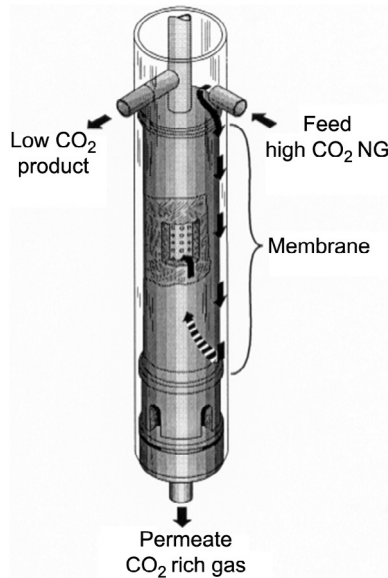


Figure 9.14 Hollow-fiber membrane element.

case, however, the central tube is for residual collection, not permeate collection.

Optimization considerations

Sleeve design

The sleeve design forces the feed to flow counter-current to the permeate instead of the more usual and less efficient co-current flow pattern.

Adjusting fiber diameters

Finer fibers give higher packing density, but larger fibers have lower permeate pressure drops and so they use the pressure driving force more efficiently.

9.9.1.6.5 Spiral Wound Versus Hollow Fiber

Each type of membrane has its own advantages and limitations.

9.9.1.6.6 Spiral Wound

Characteristics of spiral wound membranes

- Installed in horizontal vessels
- Operate at higher allowable operating pressures 1085 psig (75 barg) and thus have higher driving force available for permeation

- More resistant to fouling
- Have a long history of service in natural gas sweetening
- Perform best with colder inlet stream gas temperatures
- Do not handle varying inlet feed quality as well as hollow fiber units installed in vertical vessels
- Require extensive pretreatment equipment with high inlet stream liquid hydrocarbon loading

9.9.1.6.7 Hollow fiber

Characteristics of hollow fiber membranes

- Installed in vertical vessels
- Offer a higher packing density
- Operate at lower inlet stream pressures 580 psig (40 barg)
- Handle higher inlet stream hydrocarbon loading better than spiral wound units
- Require inlet feed gas chilling
- Hollow fiber based plants are typically smaller than spiral wound-based plants
- Handle varying inlet feed quality better than spiral wound units installed in horizontal vessels

9.9.1.6.8 Membrane Modules

Once the membranes have been manufactured into elements, they are joined together and inserted into a tube (Figure 9.15).

9.9.1.6.9 Major Vendors

Spiral wound

- UOP
- Medal
- Kvaerner

Hollow fiber

- Cynara (NATCO)

9.9.1.7 Membrane Skids

Multiple tubes are mounted on skids in either a horizontal or vertical orientation, depending on the membrane company. Figure 9.16 shows an example of a skid with horizontal tubes.

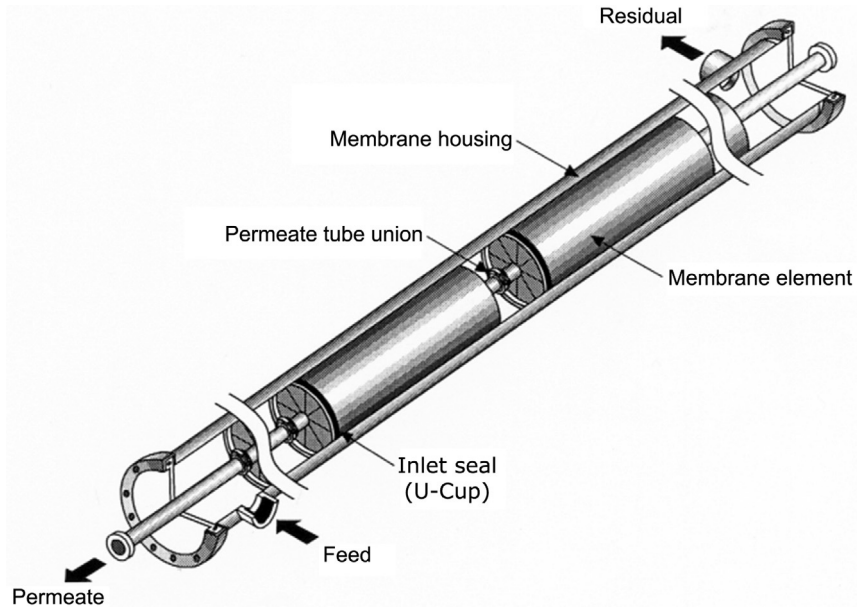


Figure 9.15 Membrane module with elements.



Figure 9.16 Horizontal membrane skid.

9.9.2 Design Considerations

9.9.2.1 Process Variables Affecting Design

9.9.2.1.1 Flow Rate

The required membrane element area is directly proportional to flow rate because the membrane systems are modular. Hydrocarbon losses (lost to

vent) are directly proportional. The percentage of hydrocarbon losses (hydrocarbon losses/feed hydrocarbons) remains the same.

9.9.2.1.2 Operating Temperature

An increase in temperature increases membrane permeability and decreases selectivity. Membrane area requirement is decreased, but the hydrocarbon losses and recycle compressor power for multistage systems are increased (Figure 9.17).

9.9.2.1.3 Feed Pressure

An increase in feed pressure decreases both membrane permeability and selectivity and creates a greater driving force across the membrane that results in a net increase in permeation through the membrane and a decrease in the membrane area requirements (Figure 9.18). Increasing the maximum operating pressure results in a less expensive and smaller system.

Limiting factors are the maximum pressure limit for the membrane elements and the cost and weight of equipment at the higher pressure rating.

9.9.2.1.4 Permeate Pressure

Exhibits the opposite effects of feed pressure

Lowens the permeate pressure

Increases the driving force, and

Lowens the membrane area requirements

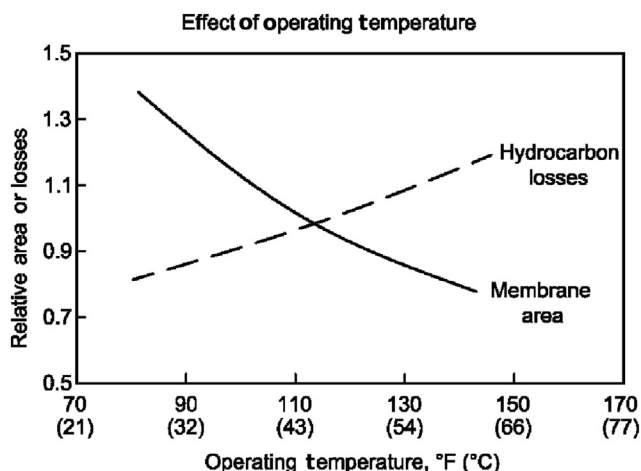


Figure 9.17 Effect of operating temperature.

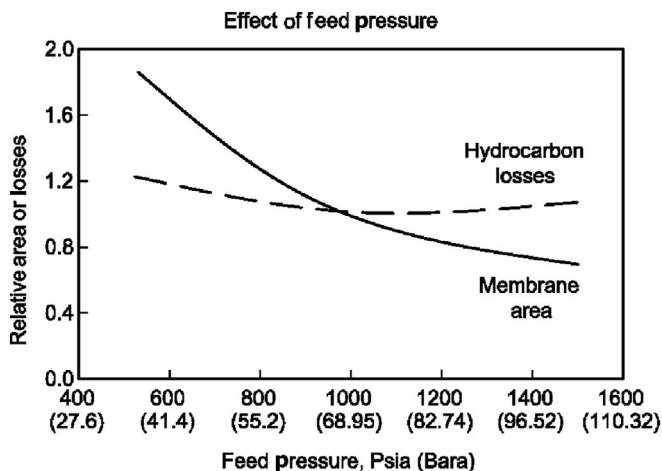


Figure 9.18 Effect of feed pressure.

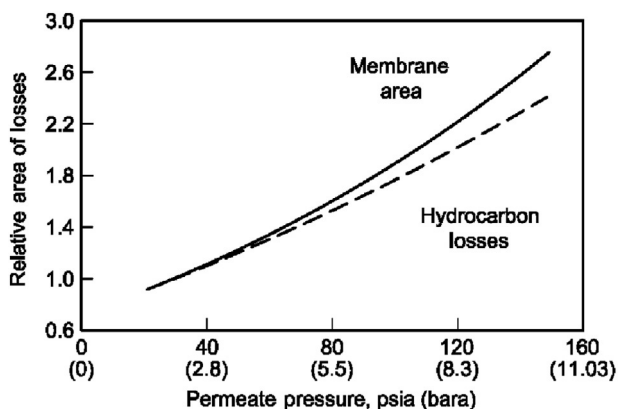


Figure 9.19 Effect of permeate pressure.

Unlike feed pressure, permeate pressure has a strong effect on hydrocarbon losses (Figure 9.19).

Pressure difference across the membrane is not the only consideration. Pressure ratio across the membrane is strongly affected by the permeate pressure.

For example,

A feed pressure of 90 bar and a permeate pressure of 3 bar produce a pressure ratio of 30.

Decreasing the permeate pressure to 1 bar increases the pressure ratio to 90 and has a dramatic effect on system performance.

Desirable to achieve the lowest possible permeate pressure

Important consideration when deciding how to further process the permeate stream

For example,

If permeate stream must be flared, then the flare design must be optimized for low pressure drop.

If permeate stream must be compressed to feed the second membrane stage or injected into a well, the increased compressor horsepower and size at lower permeate pressure must be balanced against the reduced membrane area requirements.

9.9.2.1.5 CO₂ Removal

For a constant sales gas CO₂ specification, an increase in feed CO₂ increases membrane area requirement and increases hydrocarbon losses (more CO₂ must permeate, and so more hydrocarbons permeate). This is shown in Figure 9.20.

Membrane area requirement is determined by the percentage of CO₂ removal rather than the feed or sales gas CO₂ specifications themselves.

For example,

A system for reducing a feed CO₂ content from 10% to 5% is similar in size to one reducing the feed from 50% to 25%, or one reducing a feed from 1% to 0.5%, if all have a CO₂ removal requirement of about 50%.

For a membrane system, the large difference in percent CO₂ removal (97 vs. 70%) means that the system for 0.1% sales gas is about three times

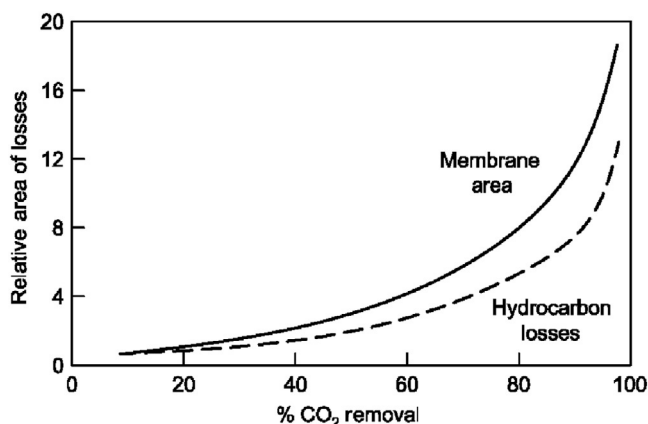


Figure 9.20 Effect of CO₂ removal.

the size of the 1% system. Traditional solvent or absorbent-based CO₂ technologies have the opposite limitation.

Their size is driven by the absolute amount of CO₂ that must be removed. For example, a system for CO₂ removal from 50% to 25% is substantially larger than one reducing CO₂ from 1% to 0.5%. For this reason, using membranes for bulk CO₂ removal and using traditional technologies for meeting low CO₂ specifications makes a lot of sense. Depending on the application, either one or both of the technologies could be used.

Changes in the feed CO₂ content of an existing membrane plant can be handled in a number of ways. Existing system can be used to produce sales gas with higher CO₂ content. An additional membrane area can be installed to meet the sales gas CO₂ content, although with increased hydrocarbon losses. Changes in the feed CO₂ content of an existing membrane plant can be handled in a number of ways. For example, if heater capacity is available, the membranes can be operated at a higher temperature to also increase capacity.

If an existing nonmembrane system must be de-bottlenecked, installing a bulk CO₂ removal system upstream of it makes good sense.

9.9.2.1.6 Other Design Considerations

9.9.2.1.7 Process Conditions

In addition to accounting for variables affecting the membrane system design, a variety of site-, country- and company-specific factors must also be considered.

9.9.2.1.8 Environmental Regulations

Environmental regulations dictate what can be done with the permeate gas, specifically whether it can be vented (cold or hot vent) to the atmosphere or flared either directly or catalytically. Ninety five to ninety nine percent CO₂ yields low Btu/scf content (flare requires a minimum of 250 Btu/scf to burn).

9.9.2.1.9 Location

Location often dictates a number of other issues, such as space and weight restrictions, level of automation, level of spares that should be available, and single versus multistage operation.

Fuel requirements can be obtained upstream of the membrane system, downstream of the pretreatment system, downstream of the membrane, and from the recycle loop in multistage systems.

Design standards, codes, and recommended practices vary from company to company. Typical areas that need to be addressed include: duplex stainless steel versus carbon steel lines, maximum pipe velocities, and painting specifications.

All items must be predetermined during the bidding stage to prevent costly modifications later.

9.9.2.2 Process Flow Schemes

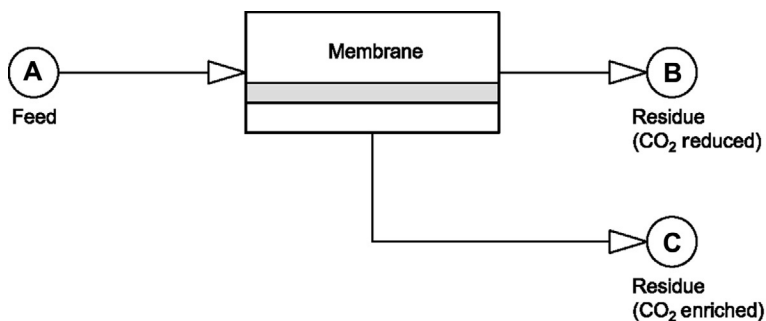
9.9.2.2.1 Single-Stage Membrane Process (Figure 9.21)

The single-stage membrane process is the simplest membrane process. Feed gas is separated into a permeate stream rich in CO_2 and a hydrocarbon rich residual stream. Methane loss is $\sim 10\%$.

9.9.2.2.2 Multistage Membrane Process

In high CO_2 removal applications, a significant amount of hydrocarbons permeate the membrane and are lost. Multistage systems attempt to recover a portion of these hydrocarbons.

The two-step membrane process allows only a portion of the first stage permeate to be lost. The portion of the first stage permeate that is lost is



	Stream		
	A	B	C
Composition (mole%)			
CH ₄	93.0	98.0	63.4
CO ₂	7.0	2.0	36.6
Flow rate (MMscfd)	20.0	17.11	2.89
Pressure (psig)	850	835	10

Methane recovery = 90.2%

Figure 9.21 Single-stage membrane process.

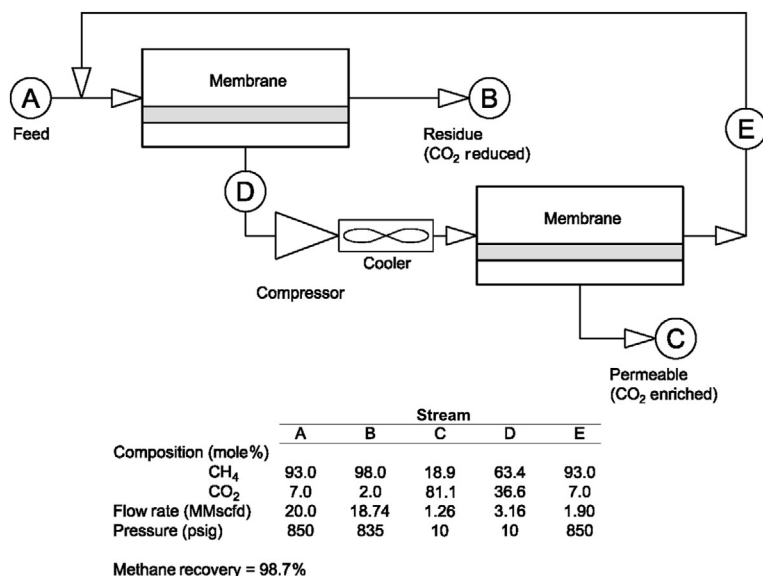


Figure 9.22 Two-stage membrane process.

usually taken from the first membrane modules, where feed CO₂, hence permeate CO₂, is highest and hydrocarbons are lowest. The remaining is recycled to the feed of the first stage. Permeate that is recycled is at low pressure and must be repressurized before it can be combined with the feed gas.

In the two-stage membrane processes (Figure 9.22), the first-stage permeate is in a second membrane stage. The permeate from the second stage, which has typically twice the CO₂ content as the first stage permeate, is vented. The residue is either recycled or combined with the feed gas. A compressor is required to repressurize the first stage permeate before it is processed in the second stage. Two-stage designs provide higher hydrocarbon recoveries than two-step or one-stage designs, but they also require more compressor horsepower because more gas must be compressed to be treated.

Other flow schemes, such as two-stage with pre-membrane flow scheme, are rarely used. Single-stage provides bulk CO₂ removal, followed by a two-stage system for final CO₂ removal. This uses a much smaller recycle compressor than that required by a standard two-stage system, although hydrocarbon losses are higher because of the single-stage portion of the system.

Many factors must be considered when deciding whether to use a single-stage or multistage system. An economic analysis must be done to ensure that

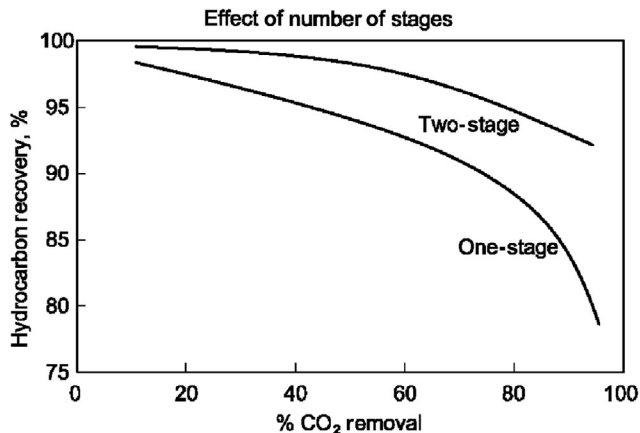


Figure 9.23 Effect of number of stages.

the cost of installing and operating a recycle compressor does not exceed the savings in hydrocarbon recovery.

Figure 9.23 plots the percentage hydrocarbon recovery versus percentage CO₂ removal for one- and two-stage systems at certain process conditions. Percentage hydrocarbon recovery is defined as the percentage of hydrocarbon recovered to the sales gas versus the hydrocarbons in the feed gas. Hydrocarbon recovery of a two-stage is significantly better than that for a single-stage system. When deciding whether to use a single- or multistage approach, one must also consider the impact of the recycle compressor. Other considerations are additional hydrocarbons used as fuel, which increases the overall hydrocarbon losses, and the significant capital cost of the compressor and maintenance. For moderate CO₂ removal applications (<50%), single-stage membrane systems usually provide better economic returns than do multistage systems.

9.9.3 Membrane Pretreatment

9.9.3.1 General Considerations

Proper pretreatment design is critical to the performance of all membrane systems. Improper pretreatment generally leads to performance decline rather than complete nonperformance.

Substances that lower performance of CO₂ removal include (1) *liquids (primarily water)* as they cause swelling and destruction of membrane

integrity; (2) **BETEX and heavy hydrocarbons (C_6 to C_{35+})** as they form a film on the membrane surface which drastically decreases permeation rate; (3) **corrosion inhibitors and well additives** as some are destructive while others are safe (one should consult with manufactures for guidance); and (4) **particulate material** as the particles can block the membrane flow area. Blockage is lower for spiral-wound than for hollow-fiber elements (low flow area). Long-term particle flow into any membrane could eventually block it.

9.9.3.2 Pretreatment System Considerations

Pretreatment must remove the aforementioned compounds. Pretreatment must also ensure that liquids will not form within the membranes themselves. Two conditions may allow condensation within the membrane. First, as the gas passes through the membrane it cools down as a result of the Joule-Thomson effect. Second, CO_2 and the lighter hydrocarbons permeate faster than the heavier hydrocarbons. The gas becomes heavier and therefore its dew point increases through the membrane. Condensation is prevented by achieving a predetermined dew point before the membrane and then heating the gas to provide a sufficient margin of superheat. Heavy hydrocarbon content can vary widely from initial prestart-up estimates and also from month to month during the plant life. Large variations are seen even between different wells in the same area. Pretreatment system must account for variations by incorporating a wide safety margin that protects the membranes against a wide range of contaminants.

9.9.3.3 Traditional Pretreatment

Figure 9.24 shows the equipment used in a traditional pretreatment system: a coalescing filter for liquids and mist elimination; a nonregenerable adsorbent guard bed for trace contaminant removal; a particle filter for dust removal after the adsorbent bed; and a heater for providing sufficient superheat to the gas.

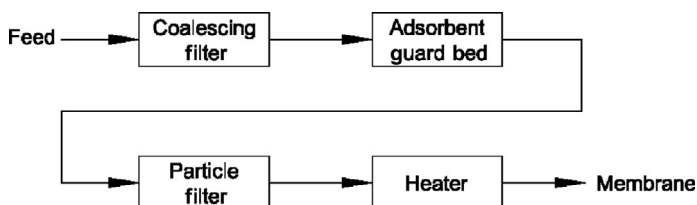


Figure 9.24 Traditional membrane pretreatment.

Figure 9.24 also shows a flow scheme that is adequate for light, stable composition gases but also has the following limitations. The adsorbent bed is the only piece of equipment that is removing heavy hydrocarbons. A sudden surge in heavy hydrocarbon content or heavier than initially estimated feed gas can saturate the adsorbent bed within days and render it useless. Since the beds are nonregenerable, they can become functional again only after the adsorbent has been replaced. Problems with the heater require that the whole membrane system be taken offline, because the heater is the only piece of equipment providing superheat.

9.9.3.4 Additions to Traditional Pretreatment

The following additional equipment may be added to enhance the system performance.

9.9.3.4.1 Chiller

A chiller may be included to reduce the dew point of the gas and the heavy hydrocarbon content. Because chilling does not completely remove all heavy hydrocarbons, an adsorbent guard bed is still required. If deep chilling is necessary, steps must be taken to prevent hydrates from forming, either by dehydrating the gas upstream or by adding hydrate inhibitors. If inhibitors are added, they may need to be removed downstream of the chiller because some inhibitors may damage the membrane.

9.9.3.4.2 Turbo-expander

The turbo-expander serves the same purpose as a chiller, but has the benefit of being a dry system. It is smaller and lighter than the refrigeration system. A disadvantage is the net pressure loss, which must be taken up by the export compressor.

9.9.3.4.3 Glycol Unit

The glycol unit is added upstream of the chiller to prevent hydrate formation or freeze-up. An adsorbent guard bed is still required to remove heavy hydrocarbons but must be larger than it would normally be because it must also remove glycol carried over from the adsorber vessels.

9.9.3.5 Enhanced Pretreatment

9.9.3.5.1 Need for Enhanced Pretreatment

It is common for an initial design, based on an extended gas analysis, to differ from actual analysis after the membrane system has been started up. For

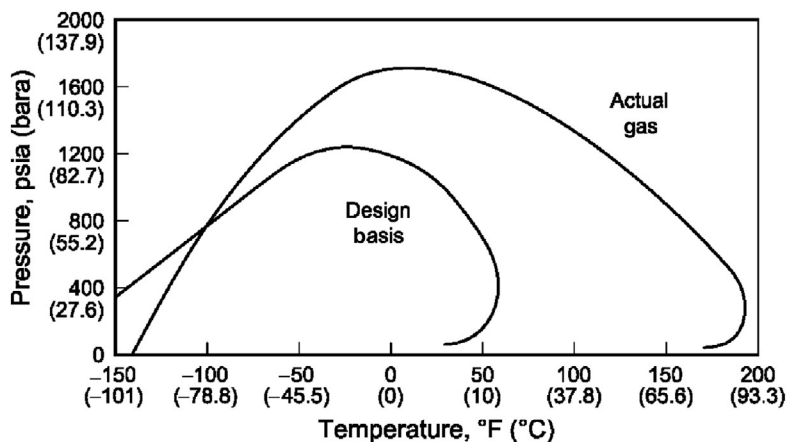


Figure 9.25 Expected and actual phase envelopes.

example, feed gas may be heavier than originally anticipated. Figure 9.25 shows the phase envelopes for the design and actual gas analysis.

Pretreatment systems may not have sufficient flexibility to handle a wide departure from design. Adsorbent beds may become fully saturated within a short time, leading to performance degradation. Preheaters may not be large enough to achieve feed temperatures that are much higher than designed. A standard way to handle a gas that is heavier than expected is to operate the membranes at a higher temperature. Temperature increase increases the margin between gas dew point and operating temperature and thus prevents condensation in the membrane.

Figure 9.26 shows an enhanced pretreatment scheme that is more suitable for cases where one or more of the following is expected: a wide variation in feed gas content; a significant amount of heavy hydrocarbons or other contaminants; or a feed gas that may be heavier than analyzed, based on the known information from nearby wells or other locations.

Feed gas is first cooled down in a heat recovery exchanger, and any condensate formed is removed in a separator and a coalescer. Liquid-free gas then enters a regenerable adsorbent guard bed system where heavy hydrocarbons and other harmful components are completely removed. Water is removed along with the heavy hydrocarbons, and thus no upstream dehydration is required. The contaminant-free gas passes through a particle filter leaving the adsorbent guard bed system. Sometimes the product gas is cooled down in a chiller whose main purpose is to reduce the hydrocarbon dew point of the feed gas. Any condensate formed in the chiller is removed in

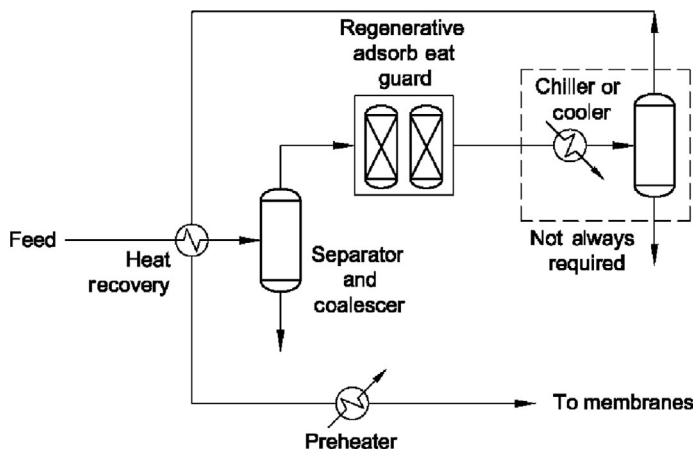


Figure 9.26 Enhanced pretreatment flow scheme.

a separator and the separator-outlet gas is routed to the feed cross exchanger. Here, the gas cools down the system feed gas and also obtains necessary superheat. Further superheat and control of membrane feed temperature are provided by a preheater.

Benefits of enhanced pretreatment follow.

9.9.3.5.2 Complete Removal of Heavy Hydrocarbons

Unlike other pretreatment schemes, the absolute cutoff of heavy hydrocarbons is possible.

9.9.3.5.3 Regenerative System

Because adsorbent guard beds are regenerable, it is better able to handle fluctuations in the heavy hydrocarbon content of the feed gas than in traditional guard beds, which require frequent replacement of adsorbent material.

9.9.3.5.4 Ability to Cope with Varying Feed Composition

Cycle time can be adjusted to provide efficient treatment of a wide variety of feed compositions and heavy hydrocarbon contents.

9.9.3.5.5 Reliability

A system can be designed to operate satisfactorily even if one of its vessels is taken offline. Critical items in the pretreatment system are usually spared so they can be serviced or maintained without shutting the system down.

9.9.3.5.6 Efficiency

A system is able to provide a number of functions, such as removal of water, heavy hydrocarbons, and mercury, that would normally be provided by separate pieces of equipment. Heat recovery is implemented in the pretreatment scheme as well as within the system itself.

9.9.4 Advantages of Membrane Systems

Membrane systems have major advantages over more traditional methods of CO₂ removal.

9.9.4.1 Lower Capital Cost Capital Expenditure (CAPEX)

Membrane systems are skid-mounted, except for larger pretreatment vessels. Scope, cost, and time required for site preparation are minimal. Installation costs are significantly lower than alternative technologies, especially for remote areas and offshore installations.

Membrane systems do not require the additional facilities, such as solvent storage and water treatment needed by other processes.

9.9.4.2 Lower Operating Costs Operating Expense (OPEX)

The only major operating cost for single-stage membrane systems is replacement. Cost is significantly lower than the solvent replacement and energy costs associated with traditional technologies. Improvements in membrane and pretreatment design allow a longer useful membrane life, which further reduces operating costs. Energy costs of multistage systems with large recycle compressors are usually comparable to those for traditional technologies.

9.9.4.3 Deferred Capital Investment

Gas flow rates often increase over time as more wells are brought online. With traditional technologies, the system design needs to take this later production into account in the initial design; thus, the majority of the equipment is installed before it is even needed. The modular nature of membrane systems means only the membranes that are needed at start-up need be installed. The rest can be added, either into existing tubes or in new skids, only when they are required. On offshore platforms, where all space requirements must be accounted for, space can be left for expansion skids rather than having to install them at the start of the project.

9.9.4.4 High Turndown

The modular nature of membrane systems means that low turndown ratios, to 10% of the design capacity or lower, can be achieved. Turnup and turn-down increments can be set at whatever level is required during the design phase.

9.9.4.5 Operational Simplicity and High Reliability

9.9.4.5.1 Single-Stage Membrane Systems

Single-stage membrane systems have no moving parts. They have almost no unscheduled downtime. They are simple to operate. They can operate unattended for long periods, provided that external upsets, such as well shut-downs, do not occur. Equipment in pretreatment systems that could cause downtime, such as filter coalescers, are usually spared so that production can continue while the equipment is under maintenance. The addition of a recycle compressor adds some complexity to the system but much less than with a solvent- or adsorbent-based technology.

9.9.4.5.2 Multistage Membrane Systems

Multistage membrane systems can be operated at full capacity as single-stage systems when the recycle compressor is down, although hydrocarbon losses will increase. Start-up, normal operation, and shutdown of a complex multistage system can be automated so that all important functions are initiated from a control room with minimal staffing.

9.9.4.6 Good Weight and Space Efficiency

Skid construction can be optimized to the space available. Multiple elements can be inserted into tubes to increase packing density. Space efficiency is especially important for offshore environments, where deck area is at a premium. [Figure 9.27](#) illustrates the space efficiency of membrane systems. The membrane unit in the lower left corner replaced all the amine and glycol plant equipment shown in the rest of the figure.

9.9.4.7 Adaptability

Because membrane area is dictated by the percentage of CO₂ removal rather than absolute CO₂ removal, small variations in feed CO₂ content hardly changes the sales gas CO₂ specification. For example, a system designed for 10% down to 3% CO₂ removal produces a 3.5% product from a 12% feed gas and a 5% product from a 15% feed gas. Adjusting process parameters such as operating temperature, the designer can further reduce the sales gas CO₂ content.

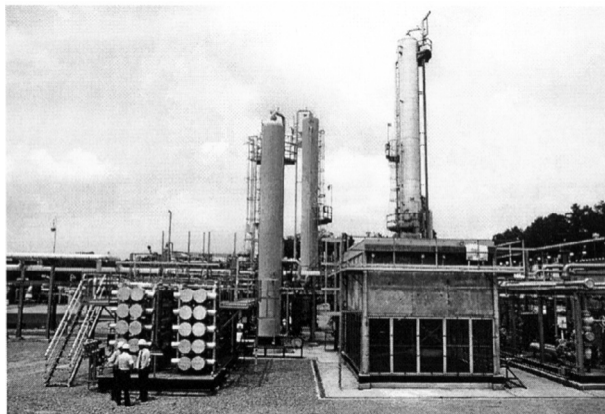


Figure 9.27 Size comparison of membrane and amine systems.

9.9.4.8 Environmental Friendly

Membrane systems do not involve the periodic removal and handling of spent solvents or adsorbents. Permeate gases can be flared, vented, or reinjected into the well or used as fuel. Items that do not need disposal, such as spent membrane elements, can be incinerated.

9.9.4.9 Design Efficiency

Membrane and pretreatment systems integrate a number of operations such as dehydration, CO₂ and H₂S removal, dew-point control, and mercury removal. Traditional CO₂ removal technologies require all of these operations as separate processes and may also require additional dehydration because some technologies saturate the product stream with water.

9.9.4.10 Power Generation

Permeate gas from membrane systems can be used to provide fuel gas for power generation, either for a recycle compressor or for other equipment. This virtually free fuel production is especially useful in membrane-amine hybrid systems, where the membrane system provides all the energy needs of the amine system.

9.9.4.11 Ideal for De-bottlenecking

Because expanding solvent- or adsorbent-based CO₂ removal plants without adding additional trains is difficult, an ideal solution is to use a membrane for bulk acid gas removal and leave the existing plant for final cleanup. An additional advantage is that the permeate gas from the membrane system can

often be based as fuel for the existing plant, thus avoiding significant increase in hydrocarbon losses.

9.9.4.12 Ideal for Remote Locations

Many of the factors mentioned above make membrane systems a highly desirable technology for remote locations where spare parts are rare and labor unskilled. Solvent storage and trucking, water supply, power generation (unless a multistage system is installed), or extensive infrastructure are not required.

9.9.5 Process Selection

9.9.5.1 Inlet Gas Stream Analysis

An accurate analysis cannot be overstressed. Process selection and economics depend on knowing all components present in the gas. Impurities, such as COS, CS₂, and mercaptans (even in small concentrations), can have a significant impact on the design of gas sweetening processes and downstream processing facilities.

When sulfur recovery is required, the composition of the acid gas stream feeding the sulfur plant must be considered. When CO₂ concentrations are >80%, selective treating should be considered to raise the H₂S concentration to the sulfur recovery unit (SRU). It may involve a multistage treating system. High concentrations of H₂O and hydrocarbons can cause design and operating problems in the SRU. The effect of these components must be weighed when selecting a gas sweetening process.

Process selection can often be based on gas composition and operating conditions. High acid gas partial pressures, 345 kPa and above, increase the likelihood of using a physical solvent. The presence of significant quantities of heavy hydrocarbons in the feed discourages the use of physical solvents. Low partial pressures of acid gases and low outlet specifications generally require the use of amines for adequate treating.

9.9.5.2 General Considerations

Each treating process has advantages relative to the others for certain applications. When making a final selection, the following facts should be considered:

- Type of acid contaminants present in the gas stream
- Concentrations of each contaminant and degree of removal required
- Volume of gas to be treated and temperature and pressure at which the gas is available

- Feasibility of recovering sulfur
- Desirability of selectively removing one or more of the contaminants without removing the others

9.9.5.3 Removal of H_2S to Meet Pipeline Qualities (4 ppm)

The presence and amount of heavy hydrocarbons and aromatics in the gas can affect the environmental conditions required at the plant site. Removal of H_2S to meet pipeline qualities can reduce the effect of these impurities at the plant.

9.9.5.3.1 Feeds with Small Acid Gas Loadings

Batch processes should be considered for feeds with small acid gas loadings. The most common processes include iron sponge, sulfa-treat and sulfa-check.

9.9.5.3.2 Feeds with Moderate to High Acid Gas Loadings

For feeds with moderate to high acid gas loadings, the disposal and replacement costs are high. There is a need to select a process that can be regenerated. Amine systems are most often used. DEA is the most commonly used amine. The acid gas stream coming off the amine stripper can be flared at moderate loadings or converted to elemental sulfur at higher loadings

9.9.5.3.3 Process Must Be Added Downstream of the Amine System

A process must be added downstream of the amine system to convert acid gas to sulfur. Processes commonly used include LO-CAT[®] and Claus. Some gas streams can be treated directly with LO-CAT[®] solution and thus, the need to separate the acid gas components from the gas stream with an amine unit is eliminated. When a Claus unit is used, it may be necessary to add tail gas cleanup downstream of the Claus unit if acid gas loadings are very high. This is normally accomplished with amine-based system because the acid gas from the stripper can be vented (assuming levels of H_2S in the gas being treated are very low). Gas permeation is attractive for low volume gas streams in remote areas where the loss of methane is not critical. Systems with a second-stage recycle may be competitive with amine systems.

9.9.5.3.4 General Considerations

Often both H_2S and CO_2 are present and must be removed. Essentially, all of the H_2S will have to be removed and only a fraction of the CO_2 will have to be removed.

9.9.5.3.5 Feeds with Low Concentrations of CO_2

For feeds with low concentrations of CO_2 , it is usually economical to use a nonselective solvent such as MEA or DEA. These processes require equipment be sized to essentially remove all the CO_2 so that the H_2S specification can be achieved.

9.9.5.3.6 Feeds with Increasing Concentrations of CO_2

For feeds with increasing concentrations of CO_2 , it is often economical to use a selective process such as MDEA, Sulfinol[®], or Selexol[®], which will remove a higher percentage of H_2S than CO_2 from a stream. Another alternative is to use gas permeation or a carbonate system for bulk removal of CO_2 upstream of a nonselective amine unit.

It may be economical to remove both H_2S and CO_2 to a level where the CO_2 content is acceptable with either a selective or nonselective process, and use a sulfur removal process (iron sponge, Sulfa-Treat, Sulfa-Check, LO-CAT[®]) for final treating of the residue gas.

9.9.5.4 Selection Charts

Figures 9.28–9.31 enable one to make a first choice of several potential candidates, which could be investigated to determine which is the most economical for a given set of conditions. These charts are not meant to replace sound engineering judgment or to cover every possible contingency.

New processes are continuously being developed. Modifications to existing proprietary products will change their range or applicability and relative cost.

9.9.5.4.1 Selection Procedure

The selection procedure requires one to

- (1) Determine the flow rate, temperature, pressure, inlet stream acid gases concentrations, and allowed concentration in the outlet stream.
- (2) Calculate the partial pressure of the acid gas components using the following equation:

$$PP_i = X_i P \quad (9.2)$$

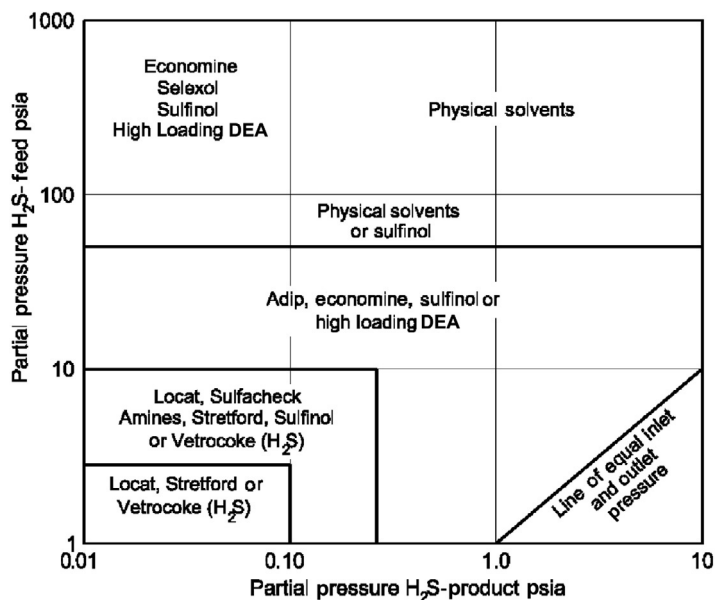


Figure 9.28 H_2S removal—no CO_2 present.

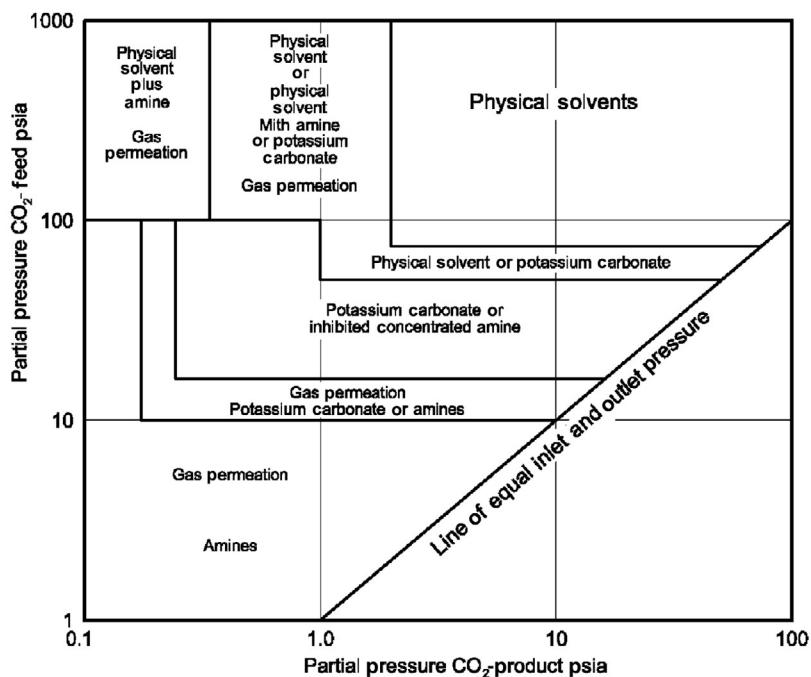


Figure 9.29 CO_2 removal—no H_2S .

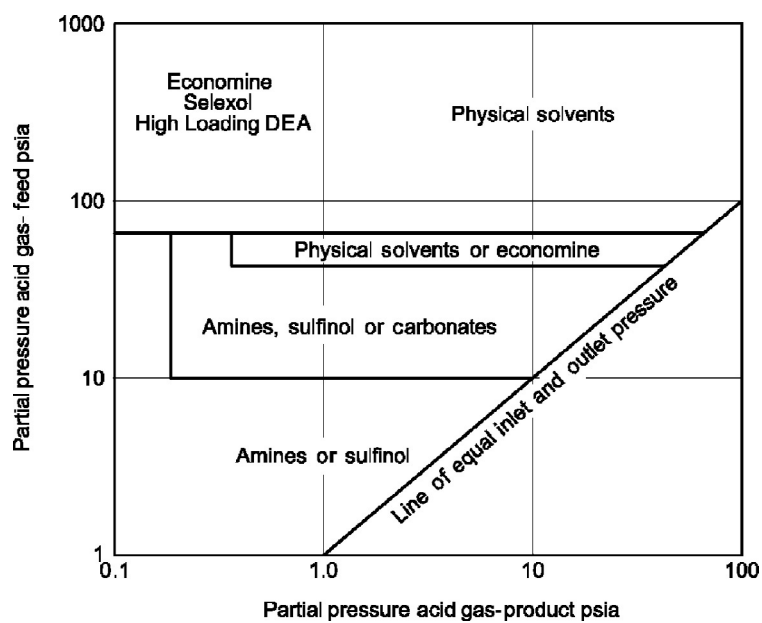


Figure 9.30 Removal of H_2S and no CO_2 .

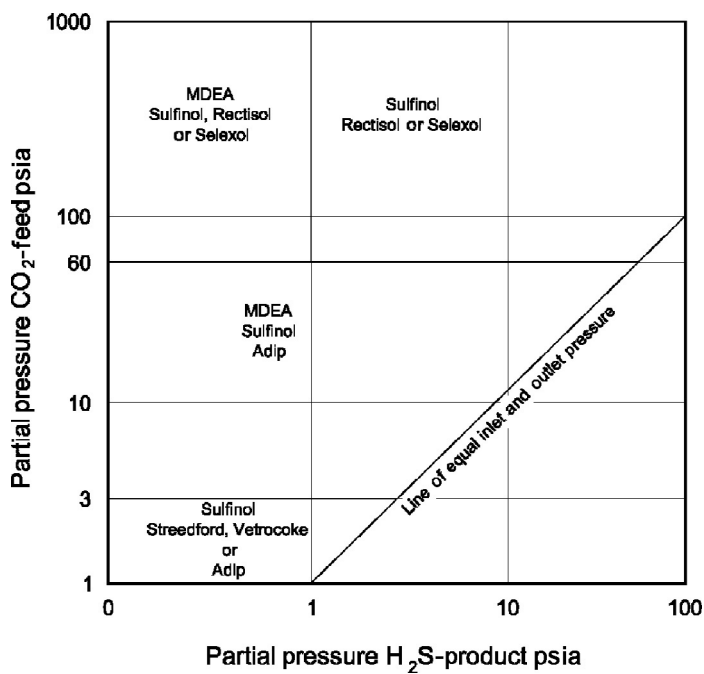


Figure 9.31 Selective removal of H_2S in the presence of CO_2 .

where PP_i = partial pressure of component i , psia (kPa); P = system pressure, psia (kPa) abs; X_i = mole fraction of component i .

Use the appropriate chart, [Figures 9.28–9.31](#), as a guide.

9.9.6 Design Procedure

9.9.6.1 Iron Sponge

9.9.6.1.1 General Considerations

The iron sponge process uses a single vessel to contain the hydrated ferric oxide wood shavings. The inlet gas line should have taps for sampling the sulfide, temperature measurement, pressure gauge and injection nozzle for methanol, water, or inhibitors.

Gas is carried into the top section of the vessel through an inlet nozzle ~ 12 in. (0.3 m) above the sponge bed. It provides uniform flow through the bed, thus minimizing the potential for channeling. Iron sponge chips are supported by a perforated heavy metal support plate and a coarse support packing material. This material may consist of scrap pipe thread protectors and 2–3 in. (50–75 mm) sections of small diameter pipe. This provides support for the bed, while offering some protection against detrimental pressure surges. Gas exits the vessel at the bottom through the vessel side-wall. This arrangement minimizes entrainment of fines. The exit line should have a pressure tap and a sample test tap.

Access to the vessel is provided through many ways in the top head and the bottom head or in the side of a larger diameter vessel (diameters > 36 in. (0.92 m) ID). For small diameter vessels, flanged top and bottom heads offer a viable solution.

The vessel is generally constructed of heat-treated carbon steel. Control of hardness is required because of the potential for sulfide stress cracking. The vessel is either internally coated, lined, or clad with stainless steel. Internal coating is typically used for vessels operating below 300 psig (2070 kPa). Cladding is used for higher operating pressures. In internally coated units, care must be exercised when setting the coarse packing to prevent damage to the coating.

9.9.6.1.2 Design Considerations

Determination of the iron sponge vessel diameter is a function of the following variables

- Desired bed life
- Velocity through the bed

- Pressure-drop through the unit
- Contact time
- Channeling potential

The following equations establish the limiting conditions for vessel sizing. Superficial gas velocity (i.e., gas flow rate divided by vessel cross-sectional area) through the iron sponge bed is normally limited to a maximum of 10 ft/s (3 m/s) at actual flow conditions to promote proper contact with the bed and prevent excessive pressure drop.

The minimum vessel diameter for gas velocity is given by
Oilfield Units

$$d_{\min} = 60 \left(\frac{Q_g T Z}{P V_{g \max}} \right)^{1/2} \quad (9.3a)$$

SI Units

$$d_{\min} = 8.58 \left(\frac{Q_g T Z}{P V_{g \max}} \right)^{1/2} \quad (9.3b)$$

where d_{\min} = minimum internal vessel diameter, cm (in.); Q_g = gas flow rate, std m³/h (MMSCFD); T = operating temperature, K (°R); Z = gas compressibility factor; P = operating pressure, kPa (psia); $V_{g \max}$ = maximum gas velocity, m/s (ft/s).

The maximum rate of deposition of 15 grains of H₂S/min-ft² (628 grains of H₂S/h m²) of bed cross-sectional area is also recommended to allow for the dissipation of the heat of reaction. The following establishes a minimum required diameter for deposition given by

Oilfield Units

$$d_{\min} = 8945 \left(\frac{Q_g \times H_2S}{\phi} \right)^{1/2} \quad (9.4a)$$

SI Units

$$d_{\min} = 4255 \left(\frac{Q_g \times H_2S}{\phi} \right)^{1/2} \quad (9.4b)$$

where ϕ = rate of deposition, grains/h m² (grains/min ft²); H₂S = mole fraction of H₂S.

The larger of the diameter calculated by Equations (9.3) or (9.4) will set the minimum vessel diameter. At very low superficial gas velocities < 2 ft/s (0.61 m/s), channeling of the gas through the bed may occur.

An upper limit to the vessel diameter may be determined by the following equation assuming a minimum velocity of 2 ft/s:

Oilfield Units

$$d_{\min} = 60 \left(\frac{Q_g TZ}{P V_{g \max}} \right)^{1/2} \quad (9.5a)$$

SI Units

$$d_{\min} = 8.58 \left(\frac{Q_g TZ}{P V_{g \max}} \right)^{1/2} \quad (9.5b)$$

where d_{\max} = maximum internal vessel diameter, cm (in.); $V_{g \min}$ = minimum gas velocity, m/s (ft/s).

A contact time of 60 s is considered a minimum in choosing a bed volume. A larger volume may be considered as it will extend the bed life and thus extend the cycle time between bed changes. Assuming a minimum contact time of 1 min, any combination of vessel diameter and bed height that satisfies the following is acceptable:

Oilfield Units

$$d^2 H \geq 3600 \frac{Q_g TZ}{P} \quad (9.6a)$$

SI Units

$$d^2 H \geq 73.63 \frac{Q_g TZ}{P} \quad (9.6b)$$

where d = vessel internal diameter, cm (in.); H = bed height, m (ft).

When selecting acceptable combinations, the bed height should be at least 10 ft (3 m) for H_2S removal and 20 ft (6 m) for mercaptan removal. The selection should produce sufficient pressure drop to ensure proper flow distribution over the entire cross-section. The vessel diameter should be between d_{\min} and d_{\max} . The iron sponge is normally sold in the United States by the bushel.

The volume in bushels can be determined from the following equation once the bed dimensions of diameter and height are known:

Oilfield Units

$$\text{Bu}_m = 7.85 \times 10^{-5} d^2 H \quad (9.7a)$$

SI Units

$$\text{Bu} = 0.0022 d^2 H \quad (9.7b)$$

where Bu = iron sponge volume, bushels; Bu_m = bed height, m (ft).

The amount of iron oxide, which is impregnated on the wood chips, is normally specified in units of pounds of iron oxide (Fe_2O_3) per bushel. Common grades are 6.5, 9, 15, or 20 lbs Fe_2O_3 /bushel.

Theoretical bed life for the iron sponge between replacements is determined from

Oilfield Units

$$t_c = 3.14 \times 10^{-8} \frac{\text{Fe } d^2 H e}{Q_g \times \text{H}_2\text{S}} \quad (9.8a)$$

SI Units

$$t_c = 1.48 \times 10^{-6} \frac{\text{Fe } d^2 H e}{Q_g \times \text{H}_2\text{S}} \quad (9.8b)$$

where t_c = cycle time, days; Fe = iron sponge content, kg $\text{Fe}_2\text{O}_3/\text{m}^3$ (lbs Fe_2O_3 /bushel); e = efficiency (0.65–0.8).

9.9.6.2 The Amine System

9.9.6.2.1 General Considerations

Equipment and the methods for designing the equipment are similar for both MEA and DEA systems. For other amine systems, the licensor should be contacted for detailed design information.

9.9.6.2.2 Amine Absorbers

Amine absorbers use counter-current flow through a trayed or packed tower to provide intimate mixing between the amine solution and the sour gas. Small diameter towers typically use stainless steel packing while larger towers

use stainless steel trays. For a system using the recommended solution concentrations and loadings, a tower with 20 actual trays is normal. Variations in solution concentrations and loadings may require further investigation to determine the number of trays.

Amine tower diameter sizing is beyond the scope of this section and is best left to the vendor. Amine absorbers for small gas flow rates commonly include an integral gas scrubber section in the bottom of the tower.

Scrubber diameter will be the same diameter as required for the absorber section. Gas leaving the scrubber will pass through a mist eliminator and then a chimney tray. The purpose of the scrubber is to remove entrained water and hydrocarbon liquids from the gas to help protect the amine.

Amine absorbers for larger gas flow rates normally have a separate scrubber vessel or microfiber filter separator so that the tower height can be decreased. This vessel should be designed according to the two-phase separator design guidelines.

For amine systems with large gas flow rates and large amine flow rates, a scrubber should be considered for the outlet sweet gas to recover carryover due to upsets or foaming. The gas leaving an amine absorber is saturated with water vapor and may require dehydration.

9.9.6.2.3 Amine Circulation Rates

The circulation rates for amine systems can be determined from the acid gas flow rates by selecting a solution concentration and an acid gas loading.

The following equations can be used:

Oilfield Units

$$L_{\text{MEA}} = \frac{112 Q_g X_A}{c_\rho A_L} \quad (9.9a)$$

SI Units

$$L_{\text{MEA}} = \frac{2.55 Q_g X_A}{c_\rho A_L} \quad (9.9b)$$

Oilfield Units

$$L_{\text{DEA}} = \frac{192 Q_g X_A}{c_\rho A_L} \quad (9.10a)$$

$$L_{\text{DEA}} = \frac{4.39 Q_g X_A}{c_\rho A_L} \quad (9.10b)$$

where L_{MEA} = MEA solution circulation rate, m^3/h (gpm); L_{DEA} = DEA solution circulation rate, m^3/h (gpm); Q_g = gas flow rate, std m^3/h (MMSCFD); X_A = required reduction in total acid gas fraction, moles acid gas removed/mole inlet gas. Note: X_A represents moles of all acid components, that is, CO_2 , H_2S , and mercaptans, as MEA and DEA are not selective; c = amine weight fraction, kg amine/kg solution (lbs amine/lbs solution); ρ = solution density, kg/m^3 (lbs/gal); A_L = acid gas loading, mole acid gas/mole amine.

The specific gravity of amine solutions at various amine concentrations and temperatures can be found in Figure 9.32 and 9.33.

For design purposes, the following solution strengths and loadings are recommended to provide an effective system without excessive corrosion rates:

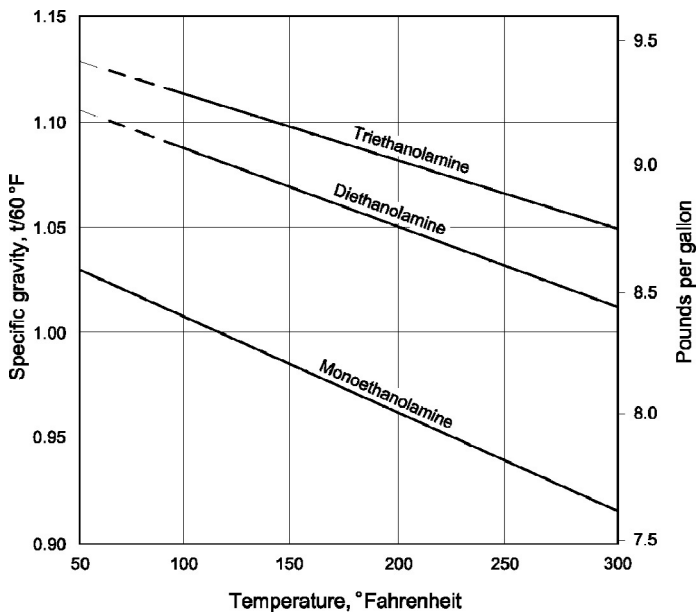


Figure 9.32 Specific gravity of amine versus temperature.

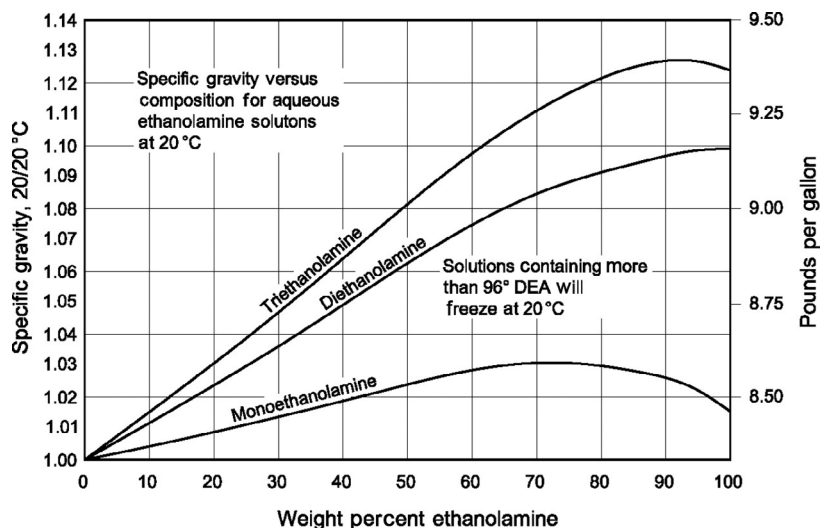


Figure 9.33 Specific gravity of amine solution versus composition. (Courtesy of Jefferson Chemicals).

MEA solution strength—20 wt% MEA

DEA solution strength—35 wt% DEA

MEA acid gas loading—0.33 mol acid gas/mol MEA

DEA acid gas loading—0.5 mol acid gas/mol DEA

Density of MEA—8.41 lbs/gal

Density of DEA—8.71 lbs/gal

Using the recommended concentrations and specific gravities at 20 °C from [Figure 9.33](#):

$$20\% \text{ MEA} = 1.008 \text{ SG} = 1.008 \times 8.34 \text{ lbs/gal}$$

$$= 8.41 \text{ lbs/gal} = 8.41 \times 0.20$$

$$= 1.68 \text{ lbs MEA/gal}$$

$$= 1.68 / 61.08 = 0.028 \text{ mol MEA/gal}$$

$$35\% \text{ DEA} = 1.044 \text{ SG} = 1.044 \times 8.34 \text{ lbs/gal}$$

$$= 8.71 \text{ lbs/gal} = 8.71 \text{ lbs/gal} \times 0.35$$

$$= 3.05 \text{ lbs DEA/gal}$$

$$= 3.05 / 105.14 = 0.029 \text{ mol DEA/gal}$$

Using these design limits, the circulation rates required can be determined from Equations (9.11) and (9.12):

Oilfield Units

$$L_{\text{MEA}} = 202 Q_g X_A \quad (9.11a)$$

SI Units

$$L_{\text{MEA}} = 0.038 Q_g X_A \quad (9.11b)$$

Oilfield Units

$$L_{\text{DEA}} = 126 Q_g X_A \quad (9.12a)$$

SI Units

$$L_{\text{DEA}} = 0.024 Q_g X_A \quad (9.12b)$$

The circulation rate determined with the above equations should be increased by 10–15% to supply an excess of amine. The rates determined can be used to size and select all equipment and piping.

9.9.6.2.4 Heat of Reaction

MEA and (MDEA) are basic solutions. These solutions react with the hydrogen sulfide and carbon dioxide to form a salt. The process of absorbing the acid gases generates heat. The heat of reaction tends to vary with the acid gas loading and the solution concentrations.

With a solution concentration from 15–25 wt% of MEA, the heat of reaction of H_2S absorbed varies from 550 to 670 Btu/lbs (1,280,000–1,558,000 J/kg), and the heat of reaction of CO_2 varies from 620 to 700 Btu/lbs (1,442,000–1,628,000 J/kg).

With a solution concentration of 25–35 wt% of DEA the heat of reaction of H_2S absorbed varies from 500 to 600 Btu/lbs (1,163,000–1,396,000 J/kg), and the heat of reaction of CO_2 varies from 580 to 650 Btu/lbs (1,349,000–1,512,000 J/kg).

Table 9.5 and Table 9.6 give the heat of reaction of CO_2 and H_2S with varying DEA solution concentrations. The heat of reaction is released when the amine and acid gas first contact and react. Thus, most of the heating takes place at the bottom of the contactor near the gas entry nozzle. As the gas goes up the tower, it exchanges heat with the amine and leaves the tower at

Table 9.5 Heat of Reaction of CO₂ in DEA Solutions
35 Weight Percent DEA

Mole Ratio CO ₂ /DEA	Joules/kg	Btu/lb CO ₂
0.2	1,730,000	744
0.4	1,479,000	636
0.5	1,310,000	563
0.6	1,140,000	490
0.8	907,000	390

25 Weight Percent DEA

Mole Ratio CO ₂ /DEA	Joules/kg	Btu/lb CO ₂
0.2	1,593,000	685
0.4	1,384,000	595
0.6	1,103,000	474
0.8	889,000	382

Table 9.6 Heat of Reaction of H₂S in DEA Solutions

Mole Ratio H ₂ S/DEA	Joules/kg	Btu/lb H ₂ S
0.2	1,405,000	604
0.3	1,342,000	577
0.4	1,279,000	550
0.6	1,177,000	506
0.8	937,000	403
1.0	484,000	208
1.2	368,000	158
1.4	323,000	139

a slightly higher temperature than the inlet amine. The inlet amine is typically 5.5 °C (10 °F) hotter than the inlet gas.

The amine outlet temperature can be estimated through a heat balance around the column where the heat into the column is the sum of the heat in the gas feed inlet, the heat in the amine inlet, and the heat due to the heat of reaction. The heat leaving the column is in the outlet gas stream, the rich amine stream, and column heat losses to the atmosphere.

9.9.6.2.5 Flash Vessel

The rich amine solution from the absorber is flashed to remove any absorbed hydrocarbons. A small amount of acid gases will also flash when the pressure

is reduced. The dissolved hydrocarbons should flash to the vapor phase and be removed. A small amount of hydrocarbon liquid may begin to collect in this separator. Thus, provisions should be made to remove these liquid hydrocarbons. Alternately, if the inlet gas to the absorber contains a high percentage of heavier hydrocarbons, a three-phase flash vessel may be installed to separate liquid hydrocarbons from the rich amine. Typically, these flash vessels provide 2–3 min of retention time for the amine solution while operating half full. Determination of the flow rate and the composition of the gas flashed in this vessel requires the use of a computer simulation program.

9.9.6.2.6 Amine Reboiler

The amine reboiler provides the heat input to an amine stripper, which reverses the chemical reactions and drives off the acid gases. Heat duty of amine reboilers varies with the system design. The higher the reboiler duty, the higher the overhead condenser duty. This gives a higher reflux ratio and thus a smaller column with fewer trays. The lower the reboiler duty, the lower the reflux ratio will be and thus the tower must have more trays.

Typically, for a stripper with 20 trays the reboiler duties will be as follows:

MEA system—1000–1200 Btu/lbs (280–330 MJ/m³) lean solution.

DEA system—900–1000 Btu/lbs (250–280 MJ/m³) lean solution.

Thus, the reboiler duty can be estimated: Equations (9.13) and (9.14)

Oilfield Units

$$q_{\text{reb}} = 72,000 L_{\text{MEA}} \quad (9.13a)$$

SI Units

$$q_{\text{reb}} = 92,905 L_{\text{MEA}} \quad (9.13b)$$

Oilfield Units

$$q_{\text{reb}} = 60,000 L_{\text{DEA}} \quad (9.14a)$$

SI Units

$$q_{\text{reb}} = 77,421 L_{\text{DEA}} \quad (9.14b)$$

where q_{reb} = reboiler duty, W (Btu/h); L_{MEA} = MEA circulation rate, m³/h (gpm); L_{DEA} = DEA circulation rate, m³/h (gpm).

Reboilers should be designed to provide the duties shown above. This will then fix the condenser duty and reflux ratio to match the stripper with 20 trays. The operating temperature for amine reboilers is determined by the operating pressure and the lean solution concentration. Typical reboiler temperature ranges are as follows:

MEA reboiler—225–260 °F (107–127 °C)

DEA reboiler—230–250 °F (110–121 °C)

For design purposes, the reboiler temperature for a stripper operating at 10 psig (69 kPa) can be assumed to be 245 °F (118 °C) for 20% MEA, and 250 °F (121 °C) for 35% DEA. Boiling point versus solution concentration curves at various pressures are shown in Figures 9.34 and 9.35.

9.9.6.2.7 Amine Stripper

Use heat from the reboiler to reverse the chemical reactions with CO₂ and H₂S and to generate steam. The steam acts as a stripping gas to remove the

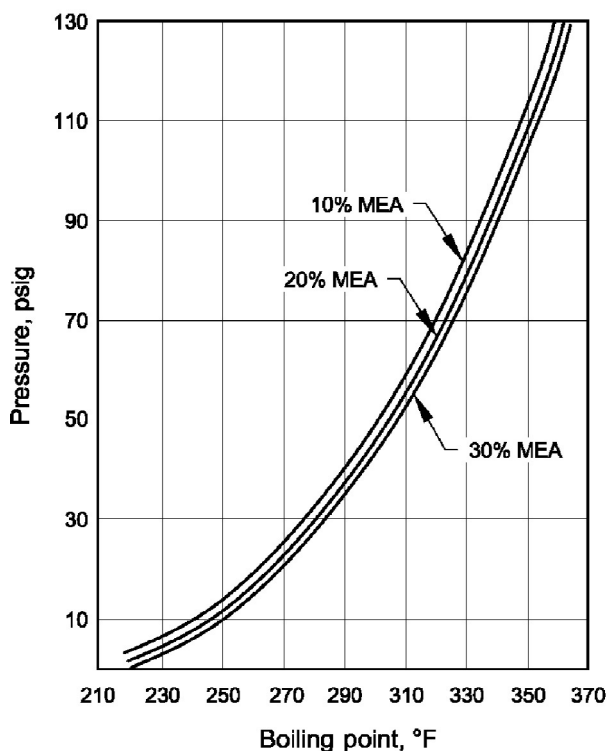


Figure 9.34 Boiling points of aqueous monoethanolamine solutions at various pressures.

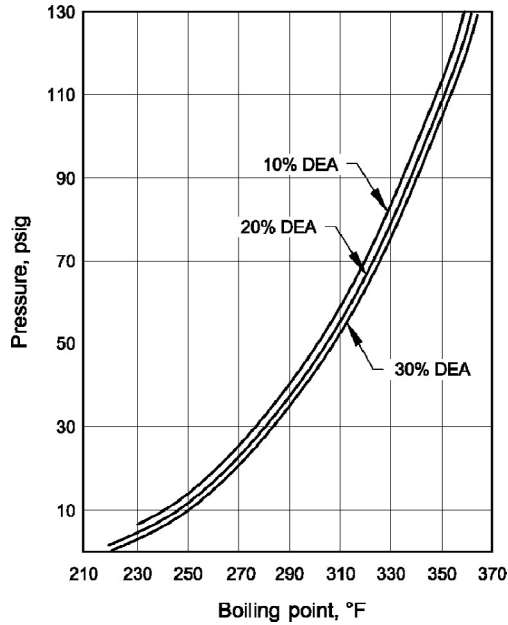


Figure 9.35 Boiling points of aqueous diethanolamine solutions at various pressures.

CO_2 and H_2S from the liquid solution and to carry these gases to the overhead. To promote mixing of the solution and the steam, the stripper is a trayed or packed tower, with packing normally used for small diameter columns and trays typically used for larger ones. A typical stripper operates at 10–15 psig (69–103 kPa) and includes 20 trays, a reboiler, and an overhead condenser.

Rich amine feed is introduced on the third/fourth tray from the top. Lean amine is removed at the bottom of the tower and acid gases are removed from the top. Maximum flow rates within the stripping column can be calculated, and then a column size can be determined using the common column sizing methods. Liquid flow rates are greatest near the bottom tray of the tower where the liquid includes the lean amine flow rate from the tower plus enough water to provide the steam generated by the reboiler.

Lean amine circulation rate is known, and from the reboiler duty, pressure, and temperature, the amount of steam generated can be estimated. Thus, the amount of water can be approximated by assuming all the heat is used to generate steam:

Oilfield Units

$$W_{\text{H}_2\text{O}} = \frac{q_{\text{reb}}}{\lambda} \quad (9.15a)$$

SI Units

$$W_{\text{H}_2\text{O}} = 3600 \frac{q_{\text{reb}}}{\lambda} \quad (9.15b)$$

where $W_{\text{H}_2\text{O}}$ = water flow rate, kg/h (lbs/h); q_{reb} = reboiler duty, w (Btu/h); λ = latent heat of vaporization of water at stripping column pressure, J/kg (Btu/lbs).

The water flow rate in gallons per minute (cubic meters per hour) is approximately:

Oilfield Units

$$L_{\text{H}_2\text{O}} = 0.002 \frac{q_{\text{reb}}}{\lambda} \quad (9.16a)$$

SI Units

$$L_{\text{H}_2\text{O}} = 3.6 \frac{q_{\text{reb}}}{\lambda} \quad (9.16b)$$

where $L_{\text{H}_2\text{O}}$ = water flow rate, m³/h (gpm).

The vapor flow rate within the tower should be calculated at both ends of the column. The higher of these vapor rates should be used in sizing the tower. At the bottom of the tower, the vapor rate equals the amount of steam generated in the reboiler (Equation 9.15). Near the top of the tower, the vapor rate equals the steam rate overhead plus the acid gas rate.

The steam rate overhead can be approximated from calculating the steam generated in the reboiler (Equation 9.15) and subtracting the amount of steam condensed by raising the amine from its inlet temperature to the reboiler temperature and the amount of steam condensed by vaporizing the acid gases.

Oilfield Units

$$W_{\text{steam}} = \frac{q_{\text{reb}} - (q_{\text{la}} - q_{\text{ra}} + q_{\text{ag}})}{\lambda} \quad (9.17a)$$

SI Units

$$W_{\text{steam}} = 3600 \frac{q_{\text{reb}} - (q_{\text{la}} - q_{\text{ra}} + q_{\text{ag}})}{\lambda} \quad (9.17b)$$

where W_{steam} = water rate overhead, kg/h (lbs/h); q_{reb} = reboiler duty, w (Btu/h); q_{la} = lean amine solution heat duty, W (Btu/h); q_{ra} = rich amine

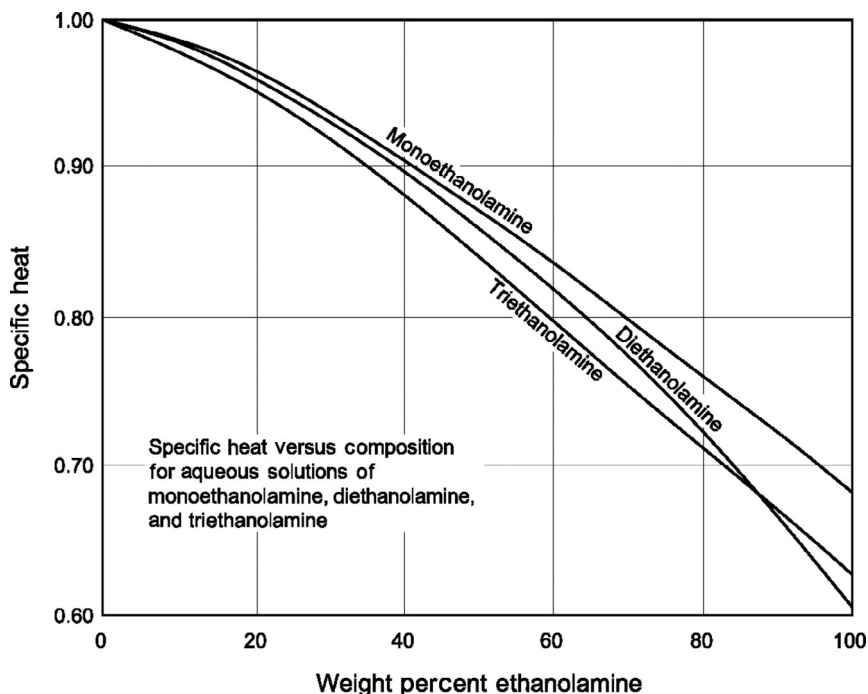


Figure 9.36 Specific heat of rich and lean amines.

solution heat duty, W (Btu/h); q_{ag} = acid gas heat duty, W (Btu/h); λ = latent heat of vaporization of water, J/kg (Btu/lbs).

The specific heat of rich amine and lean amine in Btu/lbs °F is shown in Figure 9.36.

9.9.6.2.8 Overhead Condenser and Reflux Accumulator

Amine stripper overhead condensers are typically air-cooled fin-fan exchangers. Once the reboiler duty has been specified, the overhead condenser duty can be determined from a heat balance around the stripper. A detailed computer simulation can be run to determine the heat balance and the condenser duty.

A simpler method, resulting in a slightly oversized condenser (<5%), is to assume that the condenser duty is equal to the regenerator duty minus the sum of the heat required to raise the lean amine from the stripper inlet temperature to the regenerator temperature and the heat of reaction of the acid gases. This method ignores heat in water vapor and acid gases leaving the condenser. The overhead condenser will cool the vapors leaving the top

of the stripper and condense some of the steam for reflux. The condenser outlet temperature is typically 130–145 °F (54–63 °C) depending on ambient temperature and is normally designed for an 20–30 °F (11–16 °C) approach to the maximum ambient temperature. Setting condenser outlet temperature/reflux drum pressure, (0–5 psig) less than the operating pressure of the regenerator, the amount of vapors leaving the condenser can be calculated as follows:

Oilfield Units

$$V_r = \frac{(P_R + 14.7)A_G}{(P_R + 14.7) - PP_{H_2O}} \times \frac{1}{24} \quad (9.18a)$$

SI Units

$$V_r = \frac{(P_R + 101.35)A_G}{(P_R + 101.35) - PP_{H_2O}} \times \frac{1}{24} \quad (9.18b)$$

where V_r =mole rate of vapor leaving condenser kg mol/h (lbs mol/h); P_R =reflux drum pressure, kPa (psig); A_G =mole acid gas/day, kg mol/day (lbs mol/day); PP_{H_2O} =partial pressure of water at the condenser outlet temperature, kPa abs (psia).

The amount of reflux can be determined by calculating the amount of steam condensed by the condenser. Use a top tray temperature of 210 °F (100 °C) as a first assumption, and calculate the amount of heat duty to cool the moles of vapor leaving the condenser from 210 °F (100 °C) to the condenser outlet temperature. The remaining overhead condenser duty is used in the condensation of steam. With the use of steam tables, the amount of reflux can be calculated as follows:

Oilfield Units

$$W_r = 3600 \frac{q_{\text{cond}} - q_{\text{vr}}}{h_s - h_L} \quad (9.19a)$$

SI Units

$$W_r = 3600 \frac{q_{\text{cond}} - q_{\text{vr}}}{h_s - h_L} \quad 9.19b$$

where q_{cond} =condenser duty, W (Btu/h); q_{vr} =heat duty to cool overhead vapors to condenser outlet temperature, W (Btu/h); h_s =enthalpy of steam at the top tray temperature, W (Btu/lbs); h_L =enthalpy of water at condenser outlet temperature, W (Btu/lbs); W_r =reflux rate, kg/h (lbs/h).

The amount of vapor entering the condenser is the sum of the vapor leaving the reflux accumulator and the water condensed for reflux. From these rates, the partial pressure of water in the vapor leaving can be calculated and the corresponding temperature read from the steam tables for the partial pressure of water. This is the temperature of the vapor leaving the tower. If this temperature is different from the one assumed, the calculations should be repeated until the assumed temperature and the calculated temperature are the same.

The reflux accumulator is a separator used to separate the acid gases and steam from the condensed water. The flow rate of overhead vapor and reflux calculated above can be used to size the reflux accumulator. A liquid retention time of 3 min should be adequate.

9.9.6.2.9 Rich/Lean Amine Exchangers

The rich/lean amine exchangers are usually shell-and-tube exchangers with the corrosive rich amine flowing through the tubes. The purpose of these exchangers is to reduce the regenerator duty by recovering some of the sensible heat from the lean amine. The flow rates and inlet temperatures are typically known. Thus, the outlet temperatures and duty can be determined by assuming an approach temperature for one outlet.

The closer the approach temperature selected, the greater the duty and heat recovered, but the larger and more expensive the exchanger. An approach temperature of about 30 °F (16 °C) provides an economic design, balancing the cost of the rich/lean exchanger and the regenerator to minimize the combined cost of the equipment. The exchanger duty can be estimated from:

Oilfield Units

$$q_{\text{MEA}} = 500 L_{\text{MEA}} S G_{\text{MEA}} C_{P\text{MEA}} \Delta T \quad (9.20a)$$

SI Units

$$q_{\text{MEA}} = 0.277 L_{\text{MEA}} S G_{\text{MEA}} C_{P\text{MEA}} \Delta T \quad (9.20b)$$

Oilfield Units

$$q_{\text{DEA}} = 500 L_{\text{DEA}} S G_{\text{DEA}} C_{P\text{DEA}} \Delta T \quad (9.21a)$$

SI Units

$$q_{\text{DEA}} = 0.277 L_{\text{DEA}} S G_{\text{DEA}} C_{P\text{DEA}} \Delta T \quad (9.21b)$$

where q_{MEA} = MEA exchanger duty, W (Btu/h); q_{DEA} = DEA exchanger, W (Btu/h); L_{MEA} = MEA circulation rate, m³/h (gpm); L_{DEA} = DEA

circulation rate, m^3/h (gpm); $C_{P\text{MEA}}$ = MEA specific heat, kJ/kg K (Btu/lbs $^{\circ}\text{F}$); $C_{P\text{DEA}}$ = DEA specific heat, kJ/kg K (Btu/lbs $^{\circ}\text{F}$); ΔT = amine temperature change, $^{\circ}\text{C}$ ($^{\circ}\text{F}$); SG_{MEA} = specific gravity of MEA; SG_{DEA} = specific gravity of DEA.

9.9.6.2.10 Amine Cooler

Proper control of the lean amine temperature entering the absorber is essential for the efficient operation of any amine plant. The amine cooler is typically an air cooler that lowers the lean amine temperature before it enters the absorber. The lean amine entering the absorber should be $\sim 10^{\circ}\text{F}$ (5°C) warmer than the sour gas entering the absorber. Lower amine temperatures cause the gas to cool in the absorber and thus condense hydrocarbons, which tends to cause foaming. Higher temperatures raise the amine vapor pressure, resulting in increased amine losses. Higher amine temperatures also increase the outlet gas temperature and increase the water content of the gas, increasing loading of downstream dehydration equipment.

The duty for the cooler can be calculated from the lean amine flow rate, the lean amine temperature leaving the rich/lean exchanger, and the sour gas inlet temperature as follows:

Oilfield Units

$$q_{\text{cooler}} = 500 L_{\text{LA}} \text{SG} C_{\text{PLA}} (T_{\text{out}} - T_{\text{in}}) \quad (9.22a)$$

SI Units

$$q_{\text{cooler}} = \frac{L_{\text{LA}} \text{SG} C_{\text{PLA}} (T_{\text{out}} - T_{\text{in}})}{3600} \quad (9.22b)$$

where q_{cooler} = lean amine cooler duty, W (Btu/h); SG = specific gravity of lean amine (water = 1.0); L_{LA} = lean amine circulation rate, m^3/h (gpm); C_{PLA} = specific heat of lean amine, kJ/kg K (Btu/lbs); T_{out} = amine cooler outlet temperature, = feed gas temperature + 2.2°C ($^{\circ}\text{F}$); T_{in} = amine cooler outlet temperature, = temperature out of rich/lean amine exchanger, $^{\circ}\text{C}$ ($^{\circ}\text{F}$).

9.9.6.2.11 Amine Solution Purification

Due to side reactions and/or degradation, a variety of contaminants will accumulate in an amine system. The method of removing these depends on the amine involved. In an MEA process, when the contaminants COS and CS_2 are present in the acid gas stream, an undesirable side reaction occurs, resulting in the formation of heat-stable salts. These salts should be removed from the system. For this reason, MEA systems usually include

a reclaimer. The reclaimer is a kettle type reboiler operating on a small side stream of lean solution. The temperature in the reclaimer is maintained such that the water and MEA boil to the overhead and back to the stripper, leaving the heat-stable salts in the reclaimer. Once full, the reclaimer is shut in and dumped to a waste disposal. The impurities are removed from the system, but the MEA bonded to the salts is also lost.

For DEA systems, a reclaimer is not required because the reactions with COS and CS₂ are reversed in the stripper. The small amount of degradation products from CO₂ can be removed by a carbon filter on a side stream of lean solution.

9.9.6.2.12 Amine Solution Pumps

A large portion of an amine plant's energy demand is created by the amine solution pumps. Usually, a booster pump and main circulation pump are included in the amine process. The booster pump, located downstream of the amine stripper, provides enough head, typically 25–40 psi (72–275 kPa), to flow the lean amine through the filtration equipment, the lean amine/rich amine exchanger, the amine cooler, and into the lean amine surge tank. Booster pumps are usually centrifugal type, and it is common to provide 100% spare capacity due to the service conditions. The booster pump horsepower can be estimated as follows:

Oilfield Units

$$\text{BHP} = \frac{\Delta PL_{LA}}{1714e} \quad (9.23a)$$

SI Units

$$\text{BHP} = \frac{\Delta PL_{LA}}{3598e} \quad (9.23b)$$

where BHP = pump brake horsepower, kW (HP); DP = differential pressure, kPa (psi); L_{LA} = lean amine circulation rate, m³/h (gpm); e = pump efficiency, = 0.7 for centrifugal pumps, = 0.9 for reciprocating pumps.

The amine circulation pumps take suction from the lean amine surge tank and boost the amine to the operating pressure of the absorber. Both centrifugal and positive displacement pumps are used in this application. The type selected usually depends on absorber operating pressure. A typical arrangement calls for three 50% capacity pumps to provide spare capacity. The circulation pump horsepower may be estimated using the same formula as for booster pumps, but with the pressure drop required to raise pressure from the surge tank operating pressure to the contactor operating pressure.

9.9.6.2.13 Procedure for Sizing an Amine System

The procedure for sizing an amine system is as follows:

- Determine a flash tank size.
- Determine rich/lean amine exchanger duty.
- Set the stripper overhead condenser outlet temperature based on a 20–30 °F (11–17 °C) approach to the maximum ambient temperature, and calculate the flow rates of acid gases and steam.
- Determine the condenser duty and reflux rate by performing an energy balance around the stripper.
- Size the reflux accumulator.
- Determine the lean amine cooler duty and the lean amine temperature leaving the rich/lean amine exchanger as determined earlier.
- Determine horsepower requirements for the booster pump and main circulation pump using Equation (9.23).

Design Examples (Oilfield Units)

Example Problem 1: Iron Sponge Unit

Given

$$Q_g = 2 \text{ MMSCFD}$$

$$SG = 0.6$$

$$H_2S = 19 \text{ ppm}$$

$$P = 1200 \text{ psig}$$

$$T = 100 \text{ °F}$$

Mercaptans are not present

Solution

Step 1. Calculate Minimum Vessel Diameter for Gas Velocity

(Equation 9.3a)

Oilfield Units:

$$d_{min} = 60 \left(\frac{Q_g TZ}{P V_{g \max}} \right)^{1/2}$$

where d_{min} = minimum internal vessel diameter, in; Q_g = gas flow rate, MMSCFD; T = operating temperature, °R; Z = gas compressibility factor; P = operating pressure, psia; $V_{g \max}$ = maximum gas velocity, ft/s

Oilfield Units:

$$d_{min} = 60 \left(\frac{Q_g TZ}{P V_{g \max}} \right)^{1/2}$$

$$z = 0.85$$

Use $V_{g \max} = 10 \text{ ft/s}$

$$d_{\min} = 60 \left(\frac{(2)(100 + 460)(0.85)}{(1200 + 14.7)(10)} \right)^{1/2}$$

$$d_{\min} = 16.8 \text{ in}$$

Step 2. Calculate Minimum Vessel Diameter for Deposition
(Equation 9.4a)

Oilfield Units:

$$d_{\min} = 8945 \left(\frac{Q_g X_{\text{H}_2\text{S}}}{\phi} \right)^{1/2}$$

where Φ = rate of deposition grains/min ft²; $X_{\text{H}_2\text{S}}$ = mole fraction of H₂S

$$d_{\min} = 8945 \left(\frac{Q_g X_{\text{H}_2\text{S}}}{\phi} \right)^{1/2}$$

Use a rate of deposition, of 15 grains/min-ft²

$$d_{\min} = 8945 \left(\frac{(2)(0.000019)}{15} \right)^{1/2}$$

$$d_{\min} = 14.2 \text{ in}$$

Step 3. Calculate Maximum Diameter (Equation 9.5a)

Oilfield Units:

$$d_{\max} = 60 \left(\frac{Q_g TZ}{P V_{g \min}} \right)^{1/2}$$

where d_{\max} = maximum internal vessel diameter, in;
 $V_{g \min}$ = minimum gas velocity, ft/s, = 2 ft/s

$$d_{\min} = 60 \left(\frac{(2)(100 + 460)(0.85)}{(1200 + 14.7)(2)} \right)^{1/2}$$

$$= 37.6 \text{ in.}$$

Therefore, any diameter from 16.8 to 37.6 in. is acceptable.

Step 4. Choose a Cycle Time of 1 Month or Longer (Equation 9.6a)

Oilfield Units:

$$t_c = 3.14 \times 10^{-8} \frac{\text{Fe} d^2 \text{He}}{Q_g X_{\text{H}_2\text{S}}}$$

where t_c = cycle time, days; Fe = iron sponge content, lb Fe₂O₃/bushel; e = efficiency (0.65 to 0.8)

$$d^2H = \left(\frac{t_c Q_g X_{H_2S}}{3.14 \times 10^{-8} F e e} \right)$$

$$d^2H = \frac{(30)(2)(0.000019)}{(3.14 \times 10^{-8})(9)(0.65)}$$

$$d^2H = 6206$$

d (in.)	H (ft)
18	19.15
20	15.52
22	12.82
24	10.77
30	6.90
36	4.79

An acceptable choice is a 30 in. O.D. vessel. Since t_c and e are arbitrary, a 10 ft bed is appropriate.

Step 5. Calculate Volume of Iron Sponge to Purchase (Equation 9.7a)
Oilfield Units:

$$Bu = 0.0044 d^2 H$$

where Bu = iron sponge volume, bushels; $Bu = 0.0044 d^2 H$;
 $Bu = 0.0044(30)^2 (10)$; $Bu = 39.6$ bushels

Example Problem 2: Amine Processing Unit (DEA)

Given

Gas volume = 100 MMSCFD
 Gas gravity = 0.67 SG (air = 1.0)
 Pressure = 1000 psig
 Gas temperature = 100 °F
 CO₂ inlet = 4.03%
 CO₂ outlet = 2%
 H₂S inlet = 19 ppm = 0.0019%
 H₂S outlet = 4 ppm
 Max. ambient temp. = 100 °F

Solution

Step 1. Process Selection

Total acid gas inlet = 4.03 + 0.0019 = 4.032%

Partial pressure of inlet acid gas = 1015 × (4.032/100) = 40.9 psia

Total acid gas outlet = 2.0%

Partial pressure of outlet acid gas = $1015 \times (2.0/100) = 20.3$ psia

From Figure 9.29 (CO₂ removal, no H₂S present) for removing CO₂ and H₂S, possible processes are: Amines, Sulfinol[®], and Carbonates.

Step 2. DEA Circulation Rate

Determine the circulation rate (Equation 9.10a):

Oilfield Units:

$$L_{\text{DEA}} = \frac{192 Q_g X_A}{c \rho A_L}$$

where L_{DEA} = DEA solution circulation rate, m³/h (gpm); Q_g = gas flow rate, std m³/h (MMSCFD); X_A = required reduction in total acid gas fraction, moles acid gas removed/mole inlet gas. Note: X_A represents moles of all acid components, i.e., CO₂, H₂S, and mercaptans, as MEA and DEA are not selective; c = amine weight fraction, kg amine/kg solution (lb amine/lb solution); ρ = solution density, kg/m³ (lb/gal); A_L = acid gas loading, mole acid gas/mole amine; r = DEA density, = 8.71 lb/gal, $c = 0.35$ lb/lb; $A_L = 0.50$ mole/mole; $Q_g = 100$ MMSCFD; $X_A = 4.032\% = 0.04032$

Note: In order to meet the H₂S outlet, virtually all the CO₂ must be removed, as DEA is not selective for H₂S.

$$L_{\text{DEA}} = \frac{192(100)(0.04032)}{(0.35)(8.71)(0.50)} = 508 \text{ gpm}$$

Add 10% for safety = 560 gpm.

Step 3. Reboiler Duty

Determine the reboiler duty (Equation 9.14a):

Oilfield Units:

$$q_{\text{reb}} = 60,000 L_{\text{DEA}}$$

where q_{reb} = reboiler duty, Btu/h; L_{DEA} = DEA circulation rate (gpm)

$$\begin{aligned} q_{\text{reb}} &= 60,000(560) \\ q_{\text{reb}} &= 33.6 \text{ MMBtu/h} \end{aligned}$$

Step 4. Absorber Heat Balance

Perform a heat balance around the absorber as follows:

Set lean amine inlet temperature at 10 °F higher than inlet gas, or 110 °F.

Assume gas leaving absorber has 5 °F approach to inlet amine, or 115 °F exit temperature.

Determine heat of reaction for CO₂ absorbed using loading and circulation with safety factor considered.

$$\begin{aligned}\text{Moles CO}_2 \text{ absorbed} &= \frac{100 \text{ MMSCFD} \times 0.04032}{379.5 \text{ SCF/mole}} \\ &= 10,620 \text{ moles/day CO}_2\end{aligned}$$

moles DEA circulated = 560 gpm \times 8.71 lb/gal \times 0.35 lb/lb \times mole/105.14 lb \times 1440 min/day = 23,400 mole DEA/day

Loading moles CO₂/mole DEA = 10,620/23,400 = 0.45

From [Table 9.5](#) (Heat of Reaction of CO₂ in DEA Solutions), using 35 weight percent DEA, the heat of reaction for CO₂ is 592 Btu/lb

Heat of reaction due to the CO₂ is 592 Btu/lb \times 10,620 moles/day \times 1/24 \times 44 lbs/mole = 11.5 MMBtu/h

$$\begin{aligned}\text{Moles H}_2\text{S absorbed} &= \frac{100 \text{ MMSCFD} \times (0.000019 - 0.000004)}{379.5 \text{ SCF/mole}} \\ &= 3.95 \text{ mole/day H}_2\text{S}\end{aligned}$$

600 Btu/lb H₂S \times 3.95 \times 1/24 \times 34 lb/mole = 3360 Btu/h

This is insignificant and can be ignored.

Calculate the DEA outlet temperature at the contactor.

Moles of air per SCF = 28.96

$$\text{Gas Flow Rate} = \frac{100 \text{ MMSCFD} \times 0.67 \times 28.96}{24 \times 379.5} = 213,000 \text{ lbs/h}$$

Heat Gained by Gas Steam

$$\begin{aligned}Q &= 213,000 \text{ lbs/h} \times (115 - 100)^\circ\text{F} \times 0.65 \text{ Btu/lb}^\circ\text{F} \\ &= 2.08 \text{ MMBtu/h}\end{aligned}$$

Heat lost to atmosphere depends on atmospheric temperature, surface area, wind velocity, etc. Assume 5% of reaction heat lost to atmosphere for an uninsulated absorber.

Total heat gained in outlet amine equals:

Heat of Reaction—Heat Gained by Gas Stream—Heat lost to Atmosphere.

$$\begin{aligned}\text{Heat Gained} &= 11.5 \text{ MMBtu/h} - 2.08 - (11.5 \times 0.05) \\ &= 8.9 \text{ MMBtu/h}\end{aligned}$$

Rich Amine Outlet Temperature Equals

$$= \frac{8.9 \text{MMBtu/h}}{560 \text{gpm} \times 8.71 \text{lb/gal} \times 60 \text{min/h} \times 915 \text{Btu/lb}^\circ\text{F}} + 110^\circ\text{F} = 143^\circ\text{F}$$

Step 5. Flash Tank

Determine flash tank size

$$\text{Operating Pressure} = 150 \text{ psig}$$

$$\text{Operating Temperature} = 143^\circ\text{F}$$

$$\text{Amine Rate} = 560 \text{ gpm}$$

$$(\text{Max}) \text{ CO}_2 \text{ Flashed} = \frac{10,620 \text{ moles/day}}{1440 \text{ min/day}} \times 44 \text{ lbs/mole} = 325 \text{ lbs/min}$$

$$(\text{Max}) \text{ H}_2\text{S Absorbed} = \text{Negligible}$$

Size for 3 min retention time operating half full.

Step 6. Rich/Lean Exchanger

Determine the rich/lean amine exchanger duty:

$$\begin{aligned} \text{Lean Amine Flow} &= 560 \text{ gpm} \times 8.71 \times 60 \\ &= 293,000 \text{ lb/h} \end{aligned}$$

$$\begin{aligned} \text{Rich Amine Flow} &= 293,000 + \frac{10,620 \text{ moles CO}_2/\text{day} \times 44 \text{ lb/mole}}{24 \text{ h/day}} \\ &\quad + \frac{4 \text{ moles H}_2\text{S/day}}{24 \text{ h/day}} \times 34 \text{ lb/mole} = 312,000 \text{ lbs/h} \end{aligned}$$

Assume DEA reboiler temperature is 250°F and 30°F approach of rich amine to lean amine.

$$\text{Rich Amine Inlet Temp} = 143^\circ\text{F (from Contactor)}$$

$$\begin{aligned} \text{Rich Amine Outlet Temp} &= 250^\circ\text{F} - 30^\circ\text{F} \\ &= 220^\circ\text{F} \end{aligned}$$

Calculate lean amine outlet temperature, assuming specific heat of rich amine is the same as the specific heat of lean amine $= 0.915 \text{ Btu/lb}^\circ\text{F}$

$$T_{\text{out}} = 250^\circ - \left[(220 - 143) \times \frac{312,000}{293,000} \right] = 168^\circ\text{F}$$

$$\begin{aligned}\text{Exchanger Duty} &= 312,000 \text{ lbs/h} \times 0.915 \times (220 - 143) \\ &= 22 \text{ MMBtu/h}\end{aligned}$$

Step 7. Stripper Overhead

Set the stripper overhead condenser outlet temperature, and calculate the flow rates of acid gas and steam. Size the condenser, and determine the reflux rate. The acid gas and steam will be vented, flared, or processed further for removal of H_2S .

Set condenser temperature at 30°F above maximum atmospheric temperature, i.e., 130°F .

From steam tables the partial pressure of water at 130°F is 2.22 psia.

Stripper and reflux drum will operate at 10 psig (24.7 psia).

Calculate the vapor rate leaving the reflux condenser:

Oilfield Units:

$$V_r = \frac{(\text{PR} + 14.7)\text{AG}}{(\text{PR} + 14.7) - \text{PP}_{\text{H}_2\text{O}}} \times \frac{1}{24}$$

where V_r = mole rate of vapor leaving condenser, lb mole/h;
PR = reflux drum pressure, psig; AG = moles acid gas/day, lb mole/day; $\text{PP}_{\text{H}_2\text{O}}$ = partial pressure of water at the condenser outlet temperature, psia

$$V_r = \frac{24.7 \times (\text{moles CO}_2 + \text{moles H}_2\text{S})}{24.7 - \text{PP}_{\text{H}_2\text{O}}}$$

$$V_r = \frac{24.7 \times (10,620 + 4)}{24.7 - 2.22} = 11,700 \text{ mole/day}$$

$$V_{\text{steam}} = 11,700 - 10,624 = 1100 \text{ mole/day}$$

$$\begin{aligned}W_{\text{H}_2\text{O}} &= 1100 \text{ mole/day} \times 18 \text{ lbs/mole} = 19,800 \text{ lb/day} \\ &= 800 \text{ lb/h lost from condensor}\end{aligned}$$

Step 8. Condenser Duty and Reflux Rate

Determine the condenser duty and reflux rate by performing an energy balance around the stripper.

$$q_{\text{reb}} = q_{\text{steam}} + q_{\text{H}_2\text{S}} + q_{\text{CO}_2} + q_{\text{amine}} + q_{\text{cond}}$$

The heat required to vaporize the acid gases (reverse the reaction) is

$$q_{\text{CO}_2} = 11.5 \text{ MMBtu/h and } q_{\text{H}_2\text{S}} = 3 \text{ MMBtu/h}$$

Calculate the heat differential between the heat in and the heat out for the lean DEA

$$q_{\text{la}} = 293,000 \text{ lbs/h} \times 0.915 (250 - 220) = 8 \text{ MMBtu/h}$$

$$q_{\text{cond}} = 33.6 - 8 - 11.5 \text{ MBtu/h}$$

$$q_{\text{cond}} = 14 \text{ MMBtu/h}$$

The reflux condenser must cool the acid gas and steam from the top tray temperature to 130 °F and condense the amount required for reflux.

Assume a top tray temperature of 210 °F.

Heat required to cool the acid gas is approximately

$$\begin{aligned} q_{\text{ag}} &= \frac{10,624 \text{ mole/day}}{24 \text{ h/day}} \times 44 \times 0.65 \text{ Btu/lb}^\circ\text{F} \times (210 - 130) \\ &= 1.01 \text{ MMBtu/h} \end{aligned}$$

$$W_{\text{reflux}} = \frac{(14 - 1.01) \text{ MMBtu/h}}{(1149.7 - 180) \text{ Btu/lb}} = 13,324 \text{ lb/h}$$

Calculate the vapors leaving the top tray

$$\begin{aligned} V_{\text{top}} &= 10,624 \text{ mole/day AG} + \frac{(13,368 + 788)}{18 \text{ lb/mole}} \times 24 \\ &= 29,440 \text{ mole/day vapors steam and acid gas} \end{aligned}$$

$$\text{Moles/day water} = 29,440 - 10,624 = 18,816$$

$$\begin{aligned} \text{PP}_{\text{H}_2\text{O}} \text{ in top vapors} &= \frac{(18,816) \times 24.7 \text{ psia}}{29,440 \text{ mole/day total}} \\ &= 15.8 \text{ psia} \end{aligned}$$

This $\text{PP}_{\text{H}_2\text{O}}$ of 15.8 psia is equivalent to a temperature of about 214 °F from [Table 9.7](#).

Recalculate, assuming top tray temperature of 214 °F.

Then heat required to cool the acid gas is:

$$\begin{aligned} q_{\text{ag}} &= \frac{10,624}{24} \times 44 \times 0.65 \times (214 - 130) \\ &= 1.060 \text{ MMBtu/h} \end{aligned}$$

$$W_{\text{reflux}} = \frac{(14 - 1.06) \text{ MMBtu/h}}{1151.2 - 180} = 12,290 \text{ lb/h}$$

$$\begin{aligned} V_{\text{top}} &= 10,624 \text{ mole}/D_{\text{AG}} + \frac{(12,244 + 788)}{18} \times 24 \\ &= 28,100 \text{ mole/day} \end{aligned}$$

$$\text{mole/day water} = 28,100 - 10,624 = 17,480$$

$$\text{PP}_{\text{H}_2\text{O}} \text{ in top vapors} = \frac{17,480}{28,100} \times 24.7 = 15.4 \text{ psia}$$

This is equivalent to a top tray temperature of 214.3 °F.

Table 9.7 Properties of Dry Saturated Steam

Temp. °F t	Abs. Pres. psia P	Specific Volume, cu ft/lb			Enthalpy, Btu/lb			Entropy, Btu/lb			Temp. °F t
		SAT.LIQUID V _f	EVAP. V _{fg}	SAT.VAPOR V _g	SAT.LIQUID h _f	EVAP. h _{fg}	SAT.VAPOR h _g	SAT.LIQUID S _t	EVAP. S _{fg}	SAT.VAPOR S _g	
32	0.08854	0.01602	3306	3306	0.00	1075.8	1075.8	0.0000	2.1877	2.1877	32
35	0.09995	0.01602	2947	2947	3.02	1074.1	1077.1	0.0061	2.1709	2.1770	38
40	0.12170	0.01602	2444	2444	8.05	1071.3	1079.3	0.0162	2.1436	2.1597	40
45	0.14752	0.01602	2036.4	2036.4	13.06	1068.4	1081.5	0.0262	2.1167	2.1429	45
50	0.17811	0.01603	1703.2	1703.2	18.07	1065.6	1083.7	0.0361	2.0903	2.1264	50
60	0.2563	0.01604	1206.6	1206.7	28.06	1059.9	1088.0	0.0555	2.0393	2.0948	80
70	0.3631	0.01606	867.8	867.9	38.04	1054.3	1092.3	0.0745	1.9902	2.0647	70
80	0.5069	0.01608	633.1	633.1	48.02	1048.6	1096.6	0.0932	1.9428	2.0360	80
90	0.6982	0.01610	468.0	468.0	57.99	1042.9	1100.9	0.1115	1.8972	2.0087	90
100	0.9492	0.01613	350.3	350.4	67.97	1037.2	1105.2	0.1295	1.8531	1.9826	100
110	1.2748	0.01617	265.3	265.4	77.94	1031.6	1109.5	0.1471	1.8106	1.9577	110
120	1.6924	0.01620	203.25	203.27	87.92	1025.8	1113.7	0.1645	1.7694	1.9339	120
130	2.2225	0.01625	157.32	157.34	97.90	1020.0	1117.9	0.1816	1.7296	1.9112	130
140	2.3886	0.01629	122.99	123.01	107.89	1014.1	1122.0	0.1984	1.6910	1.8894	140
150	3.718	0.01634	97.06	97.07	117.89	1008.2	1126.1	0.2149	1.6537	1.8685	150
160	4.741	0.01639	77.27	77.29	127.89	1002.3	1130.2	0.2311	1.6174	1.8485	160
170	5.992	0.01645	62.04	62.06	137.90	996.3	1134.2	0.2472	1.5822	1.8293	170
180	7.510	0.01651	50.21	50.23	147.92	990.2	1138.1	0.2630	1.5480	1.8109	180
190	9.339	0.01657	40.94	40.96	157.95	984.1	1142.0	0.2785	1.5147	1.7932	190
200	11.526	0.01663	33.62	33.64	167.99	977.9	1145.9	0.2938	1.4824	1.7762	200
210	14.123	0.01670	27.80	27.82	178.05	971.6	1149.7	0.3090	1.4508	1.7598	210
212	14.696	0.01672	26.78	26.80	180.07	970.3	1150.4	0.3120	1.4446	1.7566	212

220	17.186	0.01677	23.13	23.15	188.13	965.2	1153.4	0.3239	1.4201	1.7440	220
230	20.780	0.01684	19.365	19.382	198.23	958.8	1157.0	0.3387	1.3901	1.7288	230
240	24.969	0.01692	16.306	16.323	208.34	952.2	1160.5	0.3531	1.3609	1.7140	240
250	29.825	0.01700	13.804	13.821	216.48	945.5	1164.0	0.3675	1.3323	1.6998	250
260	35.429	0.01709	11.746	11.763	228.64	938.7	1167.3	0.3817	1.3043	1.6860	260
270	41.858	0.01717	10.044	10.061	238.84	931.8	1170.6	0.3958	1.2769	1.6727	270
280	49.203	0.01726	8.628	8.645	249.06	924.7	1173.8	0.4096	1.2501	1.6597	280
290	57.556	0.01735	7.444	7.461	259.31	917.5	1176.8	0.4234	1.2238	1.6472	290
300	67.013	0.01745	6.449	6.466	269.59	910.1	1179.7	0.4369	1.1980	1.6350	300
310	77.68	0.01755	5.609	5.625	279.92	902.6	1182.5	0.4504	1.1727	1.6231	310
320	89.66	0.01765	4.896	4.914	290.28	894.9	1185.2	0.4637	1.1478	1.6115	320
330	103.06	0.01776	4.269	4.307	300.68	887.0	1187.7	0.4769	1.1233	1.6002	330
340	118.01	0.01787	3.770	3.788	311.13	879.0	1190.1	0.4900	1.0992	1.5891	340
350	134.63	0.01799	3.324	3.342	321.63	870.7	1192.3	0.5029	1.0754	1.5783	350
360	153.04	0.01811	2.939	2.957	332.18	862.2	1194.4	0.5158	1.0519	1.5677	360
370	173.37	0.01823	2.606	2.625	342.79	853.5	1196.3	0.5286	1.0287	1.5573	370
380	195.77	0.01836	2.317	2.335	353.45	844.6	1198.1	0.5413	1.0059	1.5471	380
390	220.37	0.01850	2.0651	2.0836	364.17	835.4	1199.6	0.5539	0.9832	1.5371	390
400	247.31	0.01864	1.8447	1.8633	374.97	826.0	1201.0	0.5664	0.9608	1.5272	400
410	276.75	0.01878	1.6512	1.6700	385.83	816.3	1202.1	0.5788	0.9386	1.5174	410
420	308.83	0.01894	1.4811	1.5000	396.77	806.3	1203.1	0.5912	0.9166	1.5078	420
430	343.72	0.01910	1.3308	1.3499	407.79	796.0	1203.8	0.6035	0.8947	1.4982	430
440	381.59	0.01926	1.1979	1.2171	418.90	785.4	1204.3	0.6158	0.8730	1.4887	440
450	422.6	0.0194	1.0799	1.0993	430.1	774.5	1204.6	0.6280	0.8513	1.4793	450
460	466.9	0.0196	0.9748	0.9944	441.4	763.2	1204.6	0.6402	0.8298	1.4700	460
470	514.7	0.0198	0.9811	0.9009	452.8	751.5	1204.3	0.6523	0.8083	1.4606	470
480	566.1	0.0200	0.7972	0.8172	464.4	739.4	1203.7	0.6645	0.7868	1.4513	480

Continued

Table 9.7 Properties of Dry Saturated Steam—cont'd

Temp. °F t	Abs. Pres. psia P	Specific Volume, cu ft/lb			Enthalpy, Btu/lb			Entropy, Btu/lb			Temp. °F t
		SAT.LIQUID V _f	EVAP. V _{fg}	SAT.VAPOR V _g	SAT.LIQUID h _f	EVAP. h _{fg}	SAT.VAPOR h _g	SAT.LIQUID S _t	EVAP. S _{fg}	SAT.VAPOR S _g	
490	621.4	0.0202	0.7221	0.7423	476.0	726.8	1202.8	0.6766	0.7653	1.4419	490
500	680.8	0.0204	0.6545	0.6749	487.8	713.9	1201.7	0.6887	0.7438	1.4325	500
520	812.4	0.0209	0.5385	0.5594	511.9	686.4	1198.2	0.7130	0.7006	1.4136	520
540	962.5	0.0215	0.4434	0.4649	536.6	656.6	1193.2	0.7374	0.6568	1.3942	540
560	1133.1	0.0221	0.3647	0.3868	562.2	624.2	1186.4	0.7621	0.6121	1.3742	560
580	1325.8	0.0228	0.2989	0.3217	588.9	588.4	1177.3	0.7872	0.5659	1.3532	580
600	1542.9	0.0236	0.2432	0.2668	617.0	548.5	1165.3	0.8131	0.5176	1.3307	600
620	1736.6	0.0247	0.1955	0.2201	646.7	503.6	1150.3	0.8398	0.4664	1.3062	620
640	2059.7	0.0260	0.1638	0.1798	678.6	452.0	1130.5	0.8679	0.4110	1.2789	640
660	2365.4	0.0278	0.1165	0.1442	714.2	390.2	1104.4	0.8987	0.3485	1.2472	660
680	2708.1	0.0305	0.0810	0.1115	757.3	309.9	1067.2	0.9351	0.2719	1.2071	680
700	3093.7	0.0369	0.0392	0.0761	823.3	172.1	995.4	0.9905	0.1484	1.1389	700
705.4	3206.2	0.0503	0.0	0.0503	902.7	0	902.7	1.0580	0	1.0580	705.4

Step 9. Reflux Accumulator

Size the reflux accumulator using the principles of two-phase separators for the following conditions:

Liquid volume = 12,290 lbs/h water

Vapor volume = 11,700 moles/D or 4.5 MMSCFD

Operating pressure = 10 psig

Operating temperature = 130 °F

Step 10. Lean Amine Cooler

Determine the lean amine cooler duty to cool the amine from the amine/amine exchanger outlet temperature of 168 °F to the contactor inlet temperature of 110 °F.

$$q = 293,000 \text{ lbs/h} \times 0.915 \times (168 - 110) \\ = 15.50 \text{ MMBtu/h}$$

Step 11. Booster Pump and Circulation Pump

Determine the lean amine booster pump and main circulation pump HP requirements.

$$\text{BHP} = \frac{(\Delta P)(L_{LA})}{1714 e}$$

Assume P of 10 psi for lean/rich exchanger, 10 psi for amine cooler, 5 psi for filter and 15 psi for associated piping

$$\Delta P = 2 \times 10 + 5 + 15 = 40 \text{ psi}$$

Assume $e = 0.65$

$$\text{BHP} = ((40)(560))/((1714)(0.65)) = 20.1 \text{ HP}$$

For circulation pump, assume $\Delta P = (1000 \text{ psi})$

$$\text{BHP} = \frac{(1000)(560)}{(1714)(0.65)} = 503 \text{ HP required for circulation pump.}$$

Design Examples (SI Units)**Example Problem 3: Iron Sponge Unit****Given**

$$Q_g = 2400 \text{ std m}^3/\text{h}$$

$$\text{SG} = 0.6$$

$$\text{H}_2\text{S} = 19 \text{ ppm}$$

$$P = 8400 \text{ kPa (A)}$$

$$T = 38 \text{ }^\circ\text{C}$$

Mercaptans are not present

Solution

Step 1. Calculate Minimum Vessel Diameter for Gas Velocity
(Equation 9.3b)

SI Units:

$$d_{\min} = 8.58 \left(\frac{Q_g T Z}{P V_{g \max}} \right)^{1/2}$$

where d_{\min} = minimum internal vessel diameter, cm; Q_g = gas flow rate, std m³/h; T = operating temperature °K; Z = gas compressibility factor, = 0.85 (Gas Processor Suppliers Association (GPSA) Fig 23-8); P = operating pressure, kPa; $V_{g \max}$ = maximum gas velocity, m/sec

Use $V_{g \max} = 3 \text{ m/s}$

$$d_{\min} = 8.58 \left[\frac{(2400)(311)(0.85)}{(8400)(3)} \right]^{1/2}$$

$$d_{\min} = 43.1 \text{ cm}$$

Step 2. Calculate Minimum Diameter for Deposition (Equation 9.4b)

SI Units:

$$d_{\min} = 4255 \left(\frac{Q_g X_{\text{H}_2\text{S}}}{\phi} \right)^{1/2}$$

where ϕ = rate of deposition grams/h m², = 628; $X_{\text{H}_2\text{S}}$ = mole fraction of H₂S, 19 ppm.

$$d_{\min} = 4255 \left(\frac{(2400)(0.000019)}{628} \right)^{1/2}$$

$$d_{\min} = 36.3 \text{ cm}$$

Step 3. Calculate Maximum Diameter (Equation 9.5b)

SI Units:

$$d_{\max} = 8.58 \left(\frac{Q_g T Z}{P V_{g \min}} \right)^{1/2}$$

where d_{\max} = maximum internal vessel diameter, cm; $V_{g \min}$ = minimum gas velocity, m/s

Use a $V_{g \min}$ of 0.61 m/s

$$d_{\max} = 8.58 \left(\frac{(2400)(311)(0.85)}{(8400)(0.61)} \right)^{1/2}$$

$$d_{\max} = 95.5 \text{ cm}$$

Therefore, any diameter from 43.1 to 95.5 cm is acceptable.

Step 4. Choose a Cycle Time of One Month or Longer (Equation 9.8b)
SI Units:

$$t_c = 1.48 \times 10^{-6} \frac{\text{Fe } d^2 H e}{Q_g X_{\text{H}_2\text{S}}}$$

where t_c = cycle time, days; Fe = iron sponge content, kg Fe_2O_3 / m^3 ; e = efficiency (0.65 to 0.8).

$$d^2 H = \frac{t_c Q_g X_{\text{H}_2\text{S}}}{1.48 \times 10^{-6} \text{Fe} e}$$

Assume Fe = 116 kg/ m^3 and efficiency = 0.65

$$d^2 H = \frac{(30)(2400)(0.000019)}{(1.48 \times 10^{-6})(116)(0.65)}$$

$$d^2 H = 12,259$$

An acceptable choice is a 76.2 cm vessel. Since t_c and e are arbitrary, a 3 m bed is appropriate.

Step 5. Calculate Volume of Iron Sponge to Purchase (Equation 9.7b)

SI Units:

$$\text{Bu} = 0.0022 d^2 H$$

$$\text{Bu}_m = 7.85 \times 10^{-5} d^2 H$$

where Bu = iron sponge volume, bushels; Bu_m = iron sponge volume, m^3 ; $\text{Bu} = 0.0022(76.2)^2 (3)$; Bu = 38 bushels

Design Examples (SI Units)

Example Problem 4: Amine Processing Unit (DEA)

Given

Gas volume = 120,000 std m^3/h

Gas gravity = 0.67 SG (air = 1.0)

Pressure = 7000 kPa (A)

Gas temperature = 38 °C

CO_2 inlet = 4.03%

CO_2 outlet = 2%

H_2S inlet = 19 ppm = 0.0019%

H_2S outlet = 4 ppm

Max. ambient temp. = 38 °C

Solution

Step 1. Process Selection

Total acid gas inlet = $4.03 + 0.0019 = 4.032\%$

Partial pressure of inlet acid gas = $7000 \times (4.032/100) = 282 \text{ kPa (41 psia)}$

Total acid gas outlet = 2.0%

Partial pressure of outlet acid gas $7000 \times (2.0/100) = 140 \text{ kPa (20 psia)}$

From Figure 9.29 (CO_2 removal, no H_2S present) for removing CO_2 and H_2S , possible processes are: Amines, Sulfinol[®], and Carbonates.

The most common selection for this application is a DEA unit.

Step 2. DEA Circulation Rate (Equation 9.8b)

Determine the circulation rate.

Oilfield Units

$d \text{ (cm)}$	$H \text{ (m)}$
55.9	3.9
60.96	3.3
66.0	3.1
71.1	2.4
76.2	2.1
91.4	1.5

$$L_{\text{DEA}} = \frac{192 Q_g X_A}{c \rho A_L}$$

where L_{MEA} = MEA circulation rate, m^3/h ; L_{DEA} = DEA circulation rate, m^3/h ; Q_g = gas flow rate, std m^3/h ; X_A = required reduction in total acid gas fraction, moles acid gas removed/mole inlet gas. Note: X_A represents moles of all acid components, i.e., CO_2 , H_2S , and mercaptans, as MEA and DEA are not selective; c = amine weight fraction, kg amine/kg solution (lb amine/lb solution); ρ = solution density, kg/m^3 (lb/gal); A_L = acid gas loading, mole acid gas/mole amine, $= 1.045 \times 1000 \text{ kg}/\text{m}^3 = 1045 \text{ kg}/\text{m}^3$; $c = 0.35 \text{ kg}/\text{kg}$; $A_L = 0.50 \text{ mole}/\text{mole}$; $Q_g = 120,000 \text{ std m}^3/\text{h}$; $X_A = 4.032\% = 0.04032$

Note: In order to meet the H_2S outlet, virtually all the CO_2 must be removed, as DEA is not selective for H_2S .

Determine the circulation rate

$$L_{\text{DEA}} = \frac{4.39(120,000)(0.04032)}{(0.35)(1045)(0.50)} = 116 \text{ m}^3/\text{h}$$

Add 10% for safety = 128 m³/h

Step 3. Reboiler Duty (Equation 9.9b)

Determine the reboiler duty

SI Units

$$q_{\text{reb}} = 77,421 L_{\text{DEA}}$$

where q_{reb} = reboiler duty, W; L_{MEA} = MEA circulation rate, m³/h (gpm); L_{DEA} = DEA circulation rate, m³/h (gpm)

$$q_{\text{reb}} = 77,421(128)$$

$$q_{\text{reb}} = 10,000,000 \text{ W}$$

Step 4. Absorber Heat Balance

Perform a heat balance around the absorber as follows:

Set lean amine inlet temperature at 5 °C higher than inlet gas, or 43 °C.

Assume gas leaving absorber has 3 °C approach to inlet amine, or 46 °C exit temperature.

Determine heat of reaction for CO₂ absorbed using loading and circulation with safety factor considered.

Moles CO₂

$$\begin{aligned} \text{moles CO}_2 \text{ absorbed} &= \frac{120,000 \text{ std m}^3/\text{h} \times 0.04032}{10.87 \text{ std m}^3/\text{mole}} \\ &= 445 \text{ moles/h CO}_2 \end{aligned}$$

$$\begin{aligned} \text{Moles DEA circulated} &= 128 \text{ m}^3/\text{h} \times 1045 \text{ kg/m}^3 \times 0.35 \text{ kg/kg} \times \text{mole}/47.7 \text{ kg} \\ &= 980 \text{ mole/h} \end{aligned}$$

$$\begin{aligned} \text{Loading moles CO}_2/\text{mole DEA} &= 445/980 \\ &= 0.45 \end{aligned}$$

From Table 9.5 (Heat of Reaction of CO₂ in DEA Solutions), using 35 weight percent DEA, the heat of reaction for CO₂ is 1,395,000 Joules/kg.

Heat of reaction due to the CO₂ is 1,395,000 Joules/kg × 445 moles/h × 1/3600 × 20 kg/mole = 3,450,000 W.

Determine heat of reaction for H₂S absorbed.

$$\begin{aligned} \text{moles H}_2\text{S absorbed} &= \frac{120,000 \text{ std m}^3/\text{h} \times (0.000019 - 0.000004)}{10.87 \text{ std m}^3/\text{mole}} \\ &= 0.166 \text{ moles h H}_2\text{S} \end{aligned}$$

Use $1,300,000 \text{ Joules/kg H}_2\text{S} \times 0.166 \text{ moles/h} \times 1/3600 \times 15.4 \text{ kg/mole} = 923 \text{ W}$

Calculate the DEA outlet temperature at the contactor.

$$\text{Gas Flow Rate} = \frac{120,000 \text{ std m}^3/\text{h} \times 0.67 \times 13.14 \text{ kg/mole}}{1087 \text{ std m}^3/\text{mole}} = 97,200 \text{ kg/h}$$

Heat Gained by Gas Stream

$$q = 97,200 \text{ kg/h} \times (46 - 38)^\circ\text{C} \times 2700 \text{ J/kg}^\circ\text{C} \times \frac{\text{h}}{3600 \text{ s}} = 580,000 \text{ W}$$

Heat lost to atmosphere depends on atmospheric temperature, surface area, wind velocity, etc. Assume 5% of reaction heat lost to atmosphere for an uninsulated absorber.

Total heat gained in outlet amine equals:

Heat of Reaction—Heat Gained by Gas Stream—Heat lost to Atmosphere.

$$\begin{aligned} \text{Heat gained} &= 3,450,000 - 580,000 - (3,450,000 \times 0.05) \\ &= 2,700,000 \text{ W} \end{aligned}$$

Rich amine outlet temperature equals

$$= \frac{(2,700,000 \text{ J/s})(3600 \text{ s/h})}{(128 \text{ m}^3/\text{h})(1045 \text{ kg/m}^3)(0.915 \text{ Btu/lb}^\circ\text{F}) \left(\frac{4187 \text{ J/kg}^\circ\text{K}}{1 \text{ Btu/lb}^\circ\text{F}} \right)} + 43^\circ\text{C}$$

Step 5. Flash Tank

Determine flash tank size using the principles on two-phase separators, based on:

Operating pressure = 1035 kPa (G)

Operating temperature = 62 °C

Amine rate = 128 m³/h

(Max) CO₂ Flashed = 445 moles/h × 20 kg/mole = 8900 kg/h

(Max) H₂S absorbed = Negligible

Size for 3 min retention time operating half full.

Step 6. Rich/Lean Exchanger

Determine the rich/lean amine exchanger duty:

$$\text{Lean Amine Flow} = 128 \text{ m}^3/\text{h} \times 1045 \text{ kg/m}^3 = 134,000 \text{ kg/h}$$

$$\begin{aligned} \text{Rich Amine Flow} &= 134,000 + 445 \text{ moles CO}_2/\text{h} \times 20 \text{ kg/mole} \\ &\quad + 0.166 \text{ moles/h} \times 15.4 \text{ kg/mole} \\ &= 143,000 \text{ kg/h} \end{aligned}$$

Assume DEA reboiler temperature is 120 and 17 °C approach of rich amine to lean amine.

Rich Amine Inlet Temperature = 62 °C (from Contactor)

Rich Amine Outlet Temperature = 120 °C – 17 °C = 103 °C

Calculate lean amine outlet temperature, assuming specific heat of rich amine is the same as the specific heat of lean amine = 3830 J/kg °K

$$T_{\text{out}} = 120^{\circ}\text{C} - \left[(103 - 62) \times \frac{143,000}{134,000} \right] = 76^{\circ}\text{C}$$

$$\begin{aligned} \text{Exchanger Duty} &= 143,000 \text{ kg/h} \times 3830 \text{ J/kg}^{\circ}\text{K} \times (103 - 62)^{\circ}\text{K} \times \frac{\text{h}}{3600 \text{ s}} \\ &= 624 \times 10^6 \text{ W} \end{aligned}$$

Step 7. Stripper Overhead

Set the stripper overhead condenser outlet temperature, and calculate the flow rates of acid gas and steam. Size the condenser, and determine the reflux rate. The acid gas and steam will be vented, flared, or processed further for removal of H₂S.

Set condenser temperature at 17 °C above maximum atmospheric temperature, i.e., 55 °C.

From steam tables the partial pressure of water at 55 °C is 15.8 kPa.

Stripper and reflux drum will operate at 170 kPa (A).

Calculate the vapor rate leaving the reflux condenser.

SI Units:

$$V_r = \frac{(P_R + 101.35)AG}{(P_R + 101.35) - PP_{\text{H}_2\text{O}}} \times \frac{1}{24}$$

where V_r = mole rate of vapor leaving condenser, kg mole/h;
PR = reflux drum pressure, kPa; AG = moles acid gas/day, kg mole/day; PP_{H₂O} = partial pressure of water at the condenser outlet temperature, kPa abs

$$V_r = \frac{170 \times (\text{moles CO}_2 + \text{moles H}_2\text{S})}{170 - PP_{\text{H}_2\text{O}}}$$

$$V_r = \frac{170 \times (445 + 0.166)}{170 - 15.8} = 491 \text{ moles/h}$$

$$V_{\text{steam}} = 491 - 445.166 = 45.6 \text{ moles/h}$$

$$\begin{aligned} W_{\text{H}_2\text{O}} &= 45.6 \text{ moles/h} \times 8.16 \text{ kg/mole} \\ &= 372 \text{ kg/h lost from condenser} \end{aligned}$$

Step 8. Condenser Duty and Reflux Rate

Determine the condenser duty and reflux rate by performing an energy balance around the stripper.

$$q_{\text{reb}} = q_{\text{steam}} + q_{\text{H}_2\text{S}} + q_{\text{CO}_2} + q_{\text{amine}} + q_{\text{cond}}$$

The heat required to vaporize the acid gases (reverse the reaction) is

$$q_{\text{CO}_2} = 3,450,000 \text{ W and } q_{\text{H}_2\text{S}}$$

Calculate the heat differential between the heat in and the heat out for the lean DEA

$$\begin{aligned} q_{\text{la}} &= 134,000 \text{ kg/h} \times 3830 \text{ J/kg}^\circ\text{K} (120 - 103) \times \frac{1 \text{ h}}{3600 \text{ s}} \\ &= 2,420,000 \text{ W} \\ q_{\text{cond}} &= 10,000,000 - 2,420,000 - 3,451,000 \\ q_{\text{cond}} &= 4,130,000 \text{ W} \end{aligned}$$

The reflux condenser must cool the acid gas and steam from the top tray temperature to 55 °C and condense the amount required for reflux.

Assume a top tray temperature of 99 °C.

Heat required to cool the acid gas is approximately

$$\begin{aligned} q_{\text{ag}} &= 445.2 \text{ moles/h} \times 20 \text{ kg/mole} \times 2700 \text{ J/kg}^\circ\text{K} \times (99 - 55) \times \frac{1}{3600} \\ &= 294,000 \text{ W} \\ W_{\text{reflux}} &= \frac{(4,130,000 - 294,000) \text{ J/s}}{2676 - 230 \text{ J/g}} = 1568 \text{ g/s} \times \frac{3600 \text{ s}}{\text{h}} \\ &= 5600 \text{ kg/h} \end{aligned}$$

Calculate the vapors leaving the top tray

$$\begin{aligned} V_{\text{top}} &= 445.2 \text{ mole/h AG} + \frac{(5600 + 372)}{8} \\ &= 1200 \text{ mole/h vapors steam and acid gas} \\ \text{PP}_{\text{H}_2\text{O}} \text{ in top vapors} &= \frac{(1200 - 445) \times 170 \text{ kPa(A)}}{1200 \text{ mole/h total}} \\ &= 107 \text{ kPa} \end{aligned}$$

This $\text{PP}_{\text{H}_2\text{O}}$ of 107 kPa is equivalent to a temperature of about 103 °C from [Table 9.6](#).

Recalculate, assuming top tray temperature of 103 °C. Then, heat required to cool the acid gas is

$$\begin{aligned} q_{\text{ag}} &= 445.2 \times 20 \times 2700 \times (103 - 55) \times \frac{1}{3600} = 321,000 \text{ W} \\ W_{\text{reflux}} &= \frac{(4,130,000 - 321,000) \text{ W} \times \frac{3600}{1000}}{2680 - 230} = 5600 \text{ kg/h} \end{aligned}$$

$$V_{\text{top}} = 445.2 \text{ mole/h} + \frac{(5600 + 372)}{8} = 1200 \text{ mole/h}$$

$$\text{mole/day water} = 1200 - 445.2 = 754.8$$

$$\text{PP}_{\text{H}_2\text{O}} \text{ in top vapors} = \frac{754.8}{1200} \times 170 = 107 \text{ kPa}$$

This is equivalent to a top tray temperature of 103 °C.

Step 9. Reflux Accumulator

Size the reflux accumulator using the principles on two-phase separators for the following conditions:

Liquid volume = 5600 kg/h water

Vapor volume = 491 moles/h or 130,000 std m³/day

Operating pressure = 170 kPa (A)

Operating temperature = 55 °C

Step 10. Lean Amine Cooler

Determine the lean amine cooler duty to cool the amine from the amine/amine exchanger outlet temperature of 76 °C to the contactor inlet temperature of 43 °C.

$$q = 134,000 \text{ kg/h} \times 3830 \text{ J/kg}^\circ\text{K} \times (76 - 43)^\circ\text{K} \times \frac{1 \text{ h}}{3600 \text{ s}}$$

$$= 4,700,000 \text{ W}$$

Step 11. Booster Pump and Circulation Pump

Determine the lean amine booster pump and main circulation pump HP requirements.

$$\text{BHP} = \frac{(\Delta P)(L_{\text{LA}})}{3598 e}$$

Assume P of 70 kPa for lean/rich exchanger, 70 kPa for amine cooler, 35 kPa for filter and 100 kPa for associated piping

$$\Delta P = 2 \times 70 + 35 + 100 = 275 \text{ kPa}$$

Assume $e = 0.65$

$$\text{BHP} = \frac{(275)(128)}{(3598)(0.65)} = 15 \text{ kW required for circulation pump.}$$

For circulation pump, assume $\Delta P = (6900 \text{ kPa})$

$$\text{BHP} = \frac{(6900)(128)}{(3598)(0.65)} = 380 \text{ kW required for circulation pump.}$$



Gas Processing



10.1 NATURAL GAS LIQUID (NGL) RECOVERY CONSIDERATIONS

10.1.1 General Considerations

“Gas processing” is used to refer to the removal of

- Ethane
- Propane
- *i*-Butane
- *n*-Butane

Liquids may be fractionated and sold as pure components, or they may be combined and sold as natural gas liquids mix., or NGLs mix.

10.1.2 Processing Objectives

The objectives of gas processing are to produce transportable gas, meet sales-gas specifications, and maximize liquid recovery. Producing transportable gas to a remote location requires the gas to be delivered without allowing condensation of hydrocarbon (HC) liquids. Condensation has two drawbacks:

1. Two-phase flow requires a larger pipe diameter than single-phase flow for the same pressure drop.
2. When the two-phase stream arrives at its destination, elaborate slug catchers may be required to remove the condensation before entering the downstream equipment.

Therefore, two alternatives exist: (1) NGL recovery at the remote site or (2) dense-fluid pipelining. To meet gas-sales specifications it is important to consider that most gas specifications contain a minimum gross heating value (GHV) specification and possibly an HC dew point requirement. If the HC liquid is more valuable as a liquid than NGL, removal should be maximized while still satisfying the minimum heating value specification. If the HC liquid is more valuable as a gas, then it may be desirable to retain it as gas, subject to the HC dew point requirement.

Table 10.1 Typical Hydrocarbon Gross Heating Values

Hydrocarbon Component	GHV/(btu/scf)
Methane	1010
Ethane	1770
Propane	2516

10.1.3 Maximizing Liquid Recovery

A normal heating value specification of about 1000 Btu/scf can be met with methane alone as is shown in Table 10.1. Gas streams containing N₂ and/or CO₂, which are incombustible, can require the presence of ethane to provide the required heating value. If the heavier HCs are more valuable as liquids, then complete liquefaction of propane and heavier HCs, and partial ethane recovery, is desirable.

Cycling of natural gas in a condensate reservoir, that is, re-injecting natural gas so as to keep the reservoir pressure above the gas dew point, will maximize ultimate NGL recovery. If the reservoir pressure is allowed to fall into the two-phase region, valuable liquids are condensed and will not be recovered.



10.2 VALUE OF NGL COMPONENTS

10.2.1 Ethane and Heavier HC Components (C₂+) can be Liquefied

Relative liquid and gas phase values of HCs are illustrated for propane. In sales gas, propane is worth the contract price of the gas, assuming it can be left in the gas and sold for its GHV.

If natural gas is worth \$5.00 per MMBtu and (Table 10.1) the heating value of propane is 2516 Btu/scf, then $(1,000,000 \text{ Btu}) \times (1 \text{ scf}/2516 \text{ Btu}) = 397.5 \text{ scf}$ propane. If propane is liquefied, the amount of liquid recovered at 60 °F is $(397.5 \text{ scf}) \times (1 \text{ gal}/36.375 \text{ scf}) = 10.9 \text{ gal}$ propane. Note: 36.375 scf/gal is read from the physical constant table under volume ratio scf gas/gal liquid.

The equivalent value of this propane as a liquid is $\$5.00/10.9 \text{ gal} = \0.459 gal^{-1} (when gas is worth $\$5.00 \text{ MMBtu}^{-1}$). If liquid propane can be sold for more than $\$0.459 \text{ gal}^{-1}$ plus the cost of liquefaction, there is an economic incentive for propane liquefaction when gas is worth $\$5.00 \text{ MMBtu}^{-1}$. Figure 10.1 shows the equivalent for NGL components as a function of gas price.

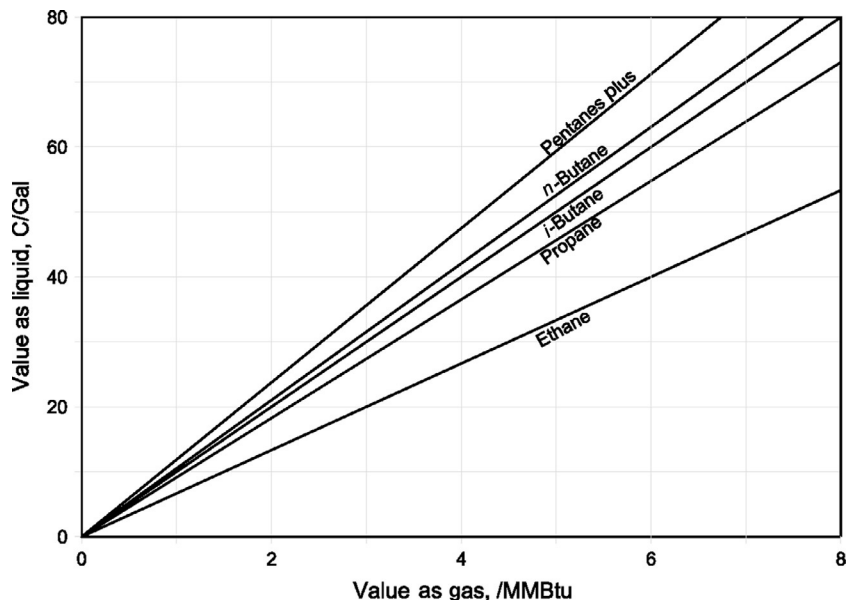


Figure 10.1 Energy equivalent of natural gas.

10.2.2 Crude Spiking

Crude spiking increases the total barrels of oil and raises the American Petroleum Institute (API) gravity (increases the sales price per barrel). Crude value is important. A value of propane is $\$0.459 \text{ gal}^{-1}$ (see above) converts to $\$23.87 \text{ bbl}^{-1}$. Crude would have to be worth more than $\$23.87 \text{ bbl}^{-1}$ to make the spiking economical. The extent of condensate removal may be limited by the sales-gas GHV specification, particularly, if appreciable N_2 and/or CO_2 are present.

Figure 10.2 shows how ethane recovery is limited by inert gas content.



10.3 GAS PROCESSING TERMINOLOGY

Definitions of LPG and NGL

Liquefied petroleum gas (LPG) products

- Defined by their vapor pressure

Unfractionated NGL

Made up of pentanes and heavier HCs

May contain some butanes and very small amounts of propane

Cannot contain heavy components that boil at more than 375°F (191°C).

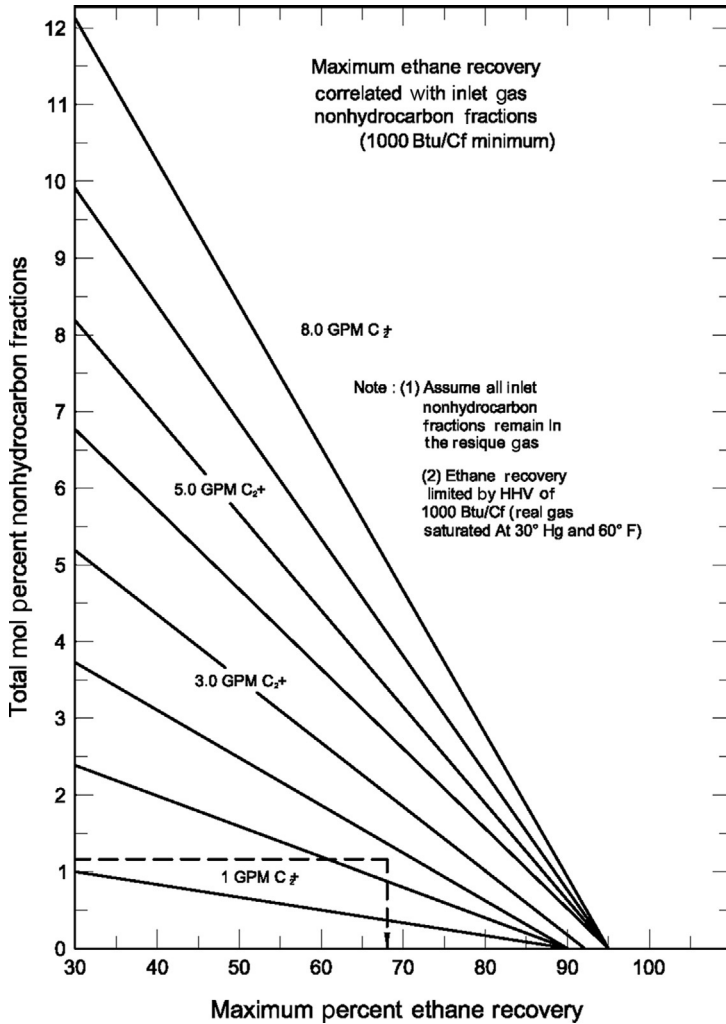


Figure 10.2 Maximum ethane recovery correlated with inlet gas nonhydrocarbon fractions.



10.4 LIQUID RECOVERY PROCESSES

10.4.1 General Considerations

Any cooling will induce condensation and yield NGL. [Figure 10.3](#) illustrates the phase diagram paths for NGL recovery. The higher the pressure, the more condensation, with other factors being equal. Another NGL recovery

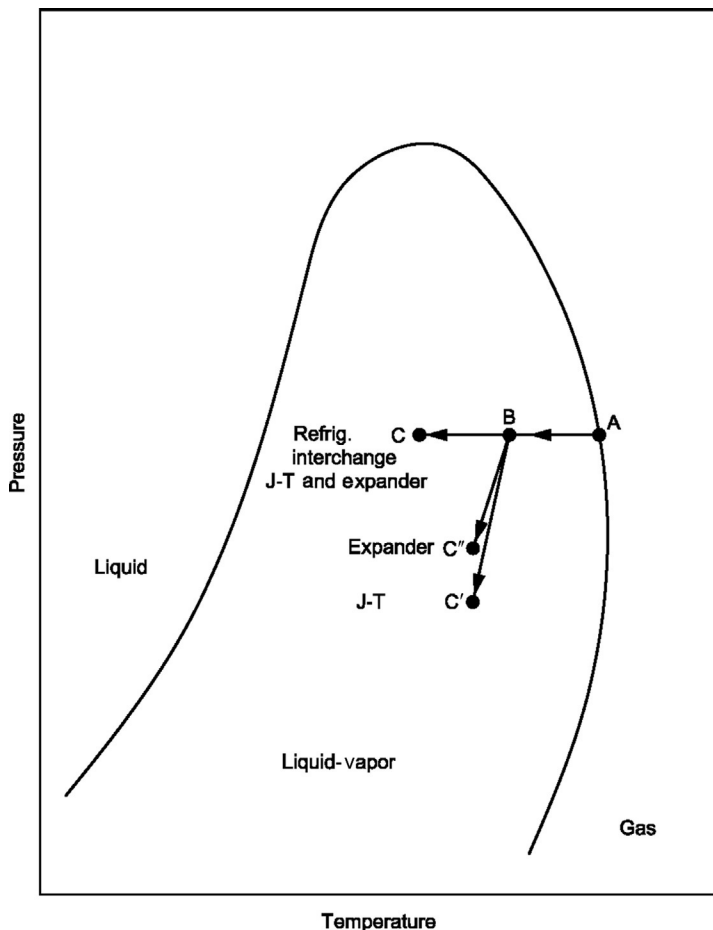


Figure 10.3 Phase diagram paths for NGL recovery.

technique is the use of a mass transfer agent. Basic NGL liquefaction processes are now described and related to [Figure 10.3](#), when possible. It is beyond the scope of this chapter to discuss detailed designs of a gas-processing plant.

10.4.2 Lean Oil Absorption Process

HC liquids can be extracted from natural gas by contacting the gas with a “lean oil,” similar to kerosene, of uniform molecular weight. The “lean oil” is used to absorb light HC compounds from the gas. The light components are separated from the rich oil and the lean oil is recycled. The fraction of each compound going into solution in the oil increases with decreasing

volatility of the compound at the absorber pressure and temperature. Although only a small percentage of the methane in the gas might go into solution, over 80% of propane, over 90% of butane, and so on, goes into solution in the oil. The lighter components, methane and ethane, are then rejected in the regeneration process of the oil, thereby capturing the absorbed propane and heavier compounds upon regeneration of the oil.

A schematic diagram of a simple lean oil absorption process is shown in Figure 10.4. The rich gas enters the absorber tower near the bottom and flows upward through the tower, which contains trays or structured packing. The absorber tower is similar in design to a glycol contact tower. Lean oil trickles down over trays or structured packing while the gas flows upward. Gas leaves the top of the absorber while the absorber oil, now rich in light HCs, leaves the bottom of the absorber tower.

As the gas flows upward in the tower, it is in intimate contact with the absorber oil, which enters the absorber tower near the top. The gas leaving the top of the tower has been stripped of most of the heavier compounds. The rich oil is then flowed to the stripper, where the oil is heated to release the absorbed HCs. The vapors leaving the top of the stripper are cooled, condensing most of the propane and heavier HCs. The vapors from the reflux separator are compressed and recycled to the rich gas or to the sales gas. The vapors from the reflux separator are compressed and recycled to the rich gas or to the sales gas.

The simple lean oil absorption processes operate at ambient temperatures. A schematic diagram of a more complex lean oil absorption process

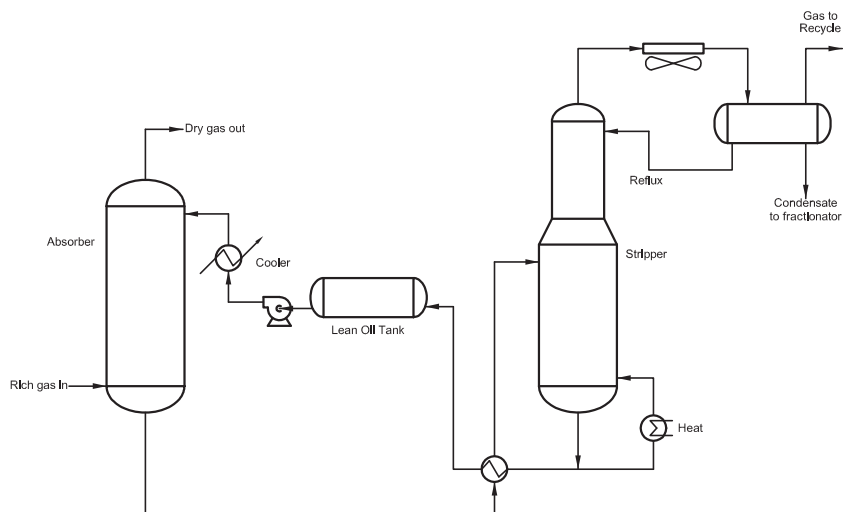


Figure 10.4 Schematic diagram of a “simple” lean oil absorption process.

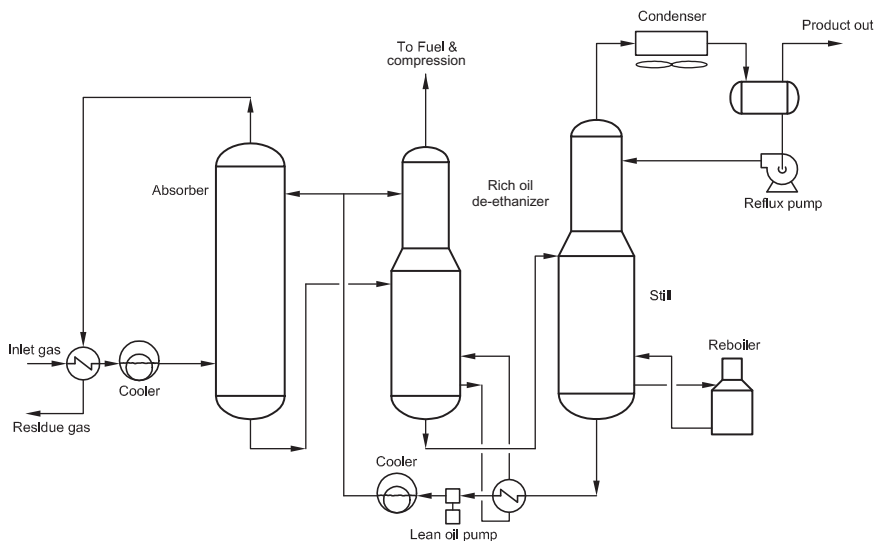


Figure 10.5 Schematic diagram of a “complex” lean oil absorption process.

is shown in [Figure 10.5](#). These processes are designed and operated at lower than ambient temperature. The inlet gas is cooled by a heat exchanger with the outlet gas and a cooler before entering the absorber.

The cooler the inlet gas stream, the higher the percentage of HCs that will be removed by the lean oil. Rich oil flows to the rich oil de-ethanizer (ROD), or de-methanizer, to reject the methane or ethane (or methane alone) as flash gas. By contacting the chilled gas from a cooler in an absorber with cooled absorption oil, more of the components in the gas go into solution than in a process operated at ambient temperatures.

In most complex lean oil plants, the rich oil dehydrator rejects both methane and ethane because very little ethane is recovered by the lean oil. If only methane were rejected, then it would be necessary to install a de-ethanizer column downstream of the still to make a separate ethane product and keep ethane from contaminating (i.e., increasing the vapor pressure of) the other liquid products made by the plant.

The ROD is similar to a cold-feed stabilizing tower for the rich oil. Heat is added at the bottom to drive off almost all of the methane (and most of the ethane) from the bottoms product by exchanging heat with the hot lean oil coming from the still. A reflux is provided by a small stream of cold lean oil injected at the top of the ROD. Gas off the tower overhead is used as plant fuel or is compressed. The amount of intermediate

components flashed with this gas can be controlled by adjusting the cold lean oil reflux rate.

Absorber oil then flows to a still where it is heated to a high enough temperature to drive off the propanes, butanes, pentanes, and other NGL components to the overhead. The closer the bottom temperature approaches the boiling temperature of the lean oil, the purer the lean oil that is recirculated to the absorber. Temperature control on the condenser keeps lean oil from being lost with the overhead.

Thus, the lean oil, in completing a cycle, goes through a recovery stage where it recovers light and intermediate components from the gas, a rejection stage where the light ends are eliminated from the rich oil, and a separation stage where the NGLs are separated from the rich oil. Because the material and heat balance calculations are rigorous, the design is normally performed by computer simulation programs.

These plants are not as popular as they once were and are rarely, if ever, constructed anymore. They are very difficult to operate, and it is difficult to predict their efficiency at removing liquids from the gas as the lean oil deteriorates with time.

Typical Liquid Recovery Levels

- $C_3 = 80\%$
- $C_4 = 90\%$
- $C_{5+} = 98\%$

10.4.3 Mechanical Refrigeration

Figure 10.6 is shows a simplified flow schematic of a mechanical refrigeration plant. Refrigeration is supplied by a vapor compression cycle. The refrigeration process is used to remove heat from a process stream. It serves a dual dew point control function and is used to meet both the HC dew point and the water dew point specification for residue or sales gas.

The minimum temperature to which the gas is cooled depends on meeting these dew point specifications. Cooling the gas lower than the minimum temperature for dew point control needs to be justified by the economics of LPG recovery. This requires comparing the value of additional LPG recovery versus the increased capital expense (CAPEX) and operating expense (OPEX). Additional recovery of LPG is achieved by chilling the gas to colder temperatures, such as -20 to -30 °F (-17 to -29 °C), or by contacting the gas stream with lean oil in an absorption tower.

Refrigeration is nothing more than transferring heat from one medium to another. Heat by itself can only flow from a higher temperature medium

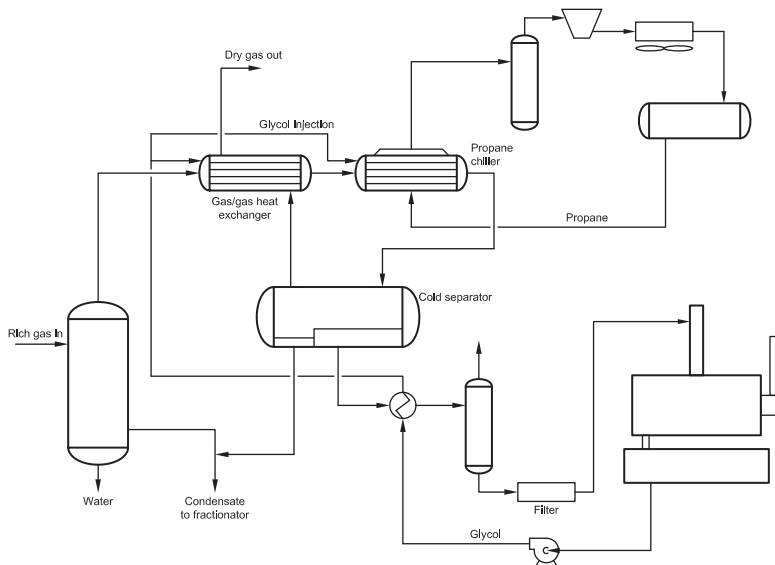


Figure 10.6 Schematic diagram of a mechanical refrigeration plant.

to a lower temperature medium. Thus, refrigeration is a process that provides the cooling medium to which the gas is exposed. Refrigeration systems normally operate trouble free but can drop in efficiency, which requires investigation.

As shown in [Figure 10.6](#), the gas/gas heat exchanger cools the inlet gas to a low enough temperature to condense the desired fraction of LPG and NGL. The refrigeration unit (propane chiller) exchanges heat with the cold gas, which has been chilled to the design cold temperature in the propane chiller. Since the gas entering the refrigeration unit is normally saturated with water vapor, and the temperature to which the gas is cooled is substantially below the hydrate formation temperature of the gas, some means of preventing hydrates must be instituted. This is normally accomplished by using hydrate inhibition, such as glycol, or by removing the water from the gas stream by using a dry desiccant such as a molecular sieve unit (see [Chapter 5](#)).

The hydrate formation temperature at a given pressure can be suppressed by the addition of a hydrate inhibitor. In conventional refrigeration units, mono-ethylene glycol (EG) is normally used because of its low cost, and at low temperatures it is not lost in the gas phase. In addition, it can be regenerated using standard regeneration units. EG is added to the natural gas stream being cooled at the following two points:

- inlet to the gas/gas heat exchanger
- inlet to the propane chiller

The refrigeration unit (propane chiller) is typically a shell-and-tube, kettle-type exchanger. Propane or some other refrigerant such as R_F-22 (which is cooled in a refrigeration cycle) is able to cool the gas to approximately -40°F . It is important to evenly distribute the glycol in the gas stream so that all gas is protected from freezing. This requires spraying the regenerated glycol evenly on to the tube sheet in these two vessels so that some glycol flows through each tube with the gas.

The refrigerant boils in the chiller at a very low, controlled temperature, removing heat from the gas stream and, thereby, condensing a portion of the gas. The cold gas, condensate, and glycol flow from the chiller to a three-phase “cold separator.” The condensate goes to a fractionation unit. The gas is sufficiently cooled so it meets both the HC and water dew points. It exchanges heat with the incoming gas to the refrigeration process.

The rich glycol is separated from the HC gas stream in the “cold separator.” The concentration of the regenerated glycol is usually on the order of 75–80% glycol with the balance being water. Sufficient glycol is injected at the two injection points to result in a mixture of water and glycol to depress the hydrate temperature to depress the hydrate formation temperature. The Hammerschmidt equation is used to determine the amount of glycol required in the water phase to lower the hydrate temperature.

Glycol and water is separated in the cold separator where they are routed to a regenerator. The water is boiled off and the glycol is circulated back to be injected into the inlet stream.

If it is desirable to recover ethane, this still is called a de-methanizer. If only propane and heavier components are to be recovered, it is called a de-ethanizer. The gas is called “plant residue” and is the outlet gas from the plant. The refrigeration process is shown as line ABC in [Figure 10.3](#). From A to B indicates gas-to-gas exchange; from B to C, chilling. Gas-to-gas exchange is very common in NGL recovery processes.

Typical Liquid Recovery Levels

- $\text{C}_3 = 85\%$
- $\text{C}_4 = 94\%$
- $\text{C}_{5+} = 98\%$

The above recovery levels are higher than those for a lean oil plant. It is possible to recover a small percentage of ethane in a refrigeration plant.

Disadvantages

- Limited by the ability to cool the inlet stream to no lower than -40°F (-40°F) with normal refrigerants.

10.4.4 Joule-Thomson Expansions

Cooling of natural gas can also be achieved by expanding high pressure gas to a lower pressure across an expansion valve or choke. This is a constant enthalpy process, and the amount of the temperature reduction depends on the pressure ratio of initial pressure divided by the final pressure, the absolute pressures and the starting temperature, as well as the gas composition. This is a practical method to cool gas and extract HC liquids if there is a lot of “free” pressure available. It is also a more practical process than the turbo-expander process for low gas rates, especially if the gas rates fluctuate.

Figure 10.7 is a simplified flow diagram of a typical Joule-Thomson (J-T) self-refrigeration process, as opposed to mechanical refrigeration. The main process equipment is the expansion valve or choke. The high-pressure gas enters through an inlet separator, which removes the condensed water and any liquid HCs. The gas streams out of the separator, then flows through a heat exchanger, exchanging heat through an inlet separator, which removes the condensed water and any liquid HCs. The gas streams out of the separator, then flows through a heat exchanger, exchanging heat with the cooled, low-pressure gas. Some water and perhaps some HC will condense in the heat exchanger from the high-pressure gas stream. The high-pressure gas then flows through the expansion valve, which drops the pressure of the gas to the design pressure. Simultaneously, a reduction in temperature occurs. Depending on the gas composition and the pressure and temperature

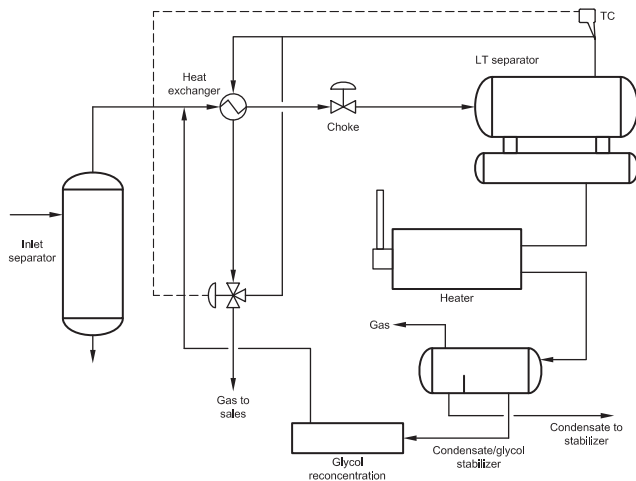


Figure 10.7 Schematic diagram of a Joule-Thomson (J-T) self-refrigeration process.

of the gas mixture, a certain amount of the mixture will condense and form a liquid HC stream. Water will condense to the equilibrium water content of the gas at the final pressure and temperature.

If the resulting temperature of the gas after the heat exchanger, or upon expansion, is below the hydrate temperature at the operating pressure, hydrates form unless the gas has been dehydrated. To avoid the formation of hydrates in water-saturated gas, a hydrate inhibitor is added to the gas stream ahead of the heat exchanger. The chemical typically used to depress the hydrate temperature is EG, but di-ethylene glycol may also be used. [Figure 10.7](#) shows the flow of the glycol and includes a reconcentration step. EG is usually regenerated to a lean concentration of about 75–80% by weight and is circulated at a rate such that the resulting final glycol concentration is sufficient to depress the hydrate forming temperature to about 5 °F below the hydrate temperature of the gas at the final pressure. The required lean glycol circulation rate is determined by the Hammerschmidt equation and depends on the water content of the gas, the concentration of the lean glycol, and the necessary hydrate temperature depression.

A bypass line around the heat exchanger on the cold gas out of the low temperature separator allows for control of the degree of cooling the process gas. To aid the separation between the cold condensate and glycol, a heater can be included in the equipment. After heating, the liquids are flowed into a three-phase separator, where the small amount of gas, the condensate, and the rich glycol are separated. The glycol is then reconcentrated with a conventional reboiler and still and is re-injected into the process gas stream.

The inlet gas passes first through the gas-to-gas exchanger and then to an expansion or “choke” valve. The expansion through the choke is essentially a constant enthalpy process. Nonideal behavior of the inlet gas causes the temperature to fall with the pressure reduction, as shown by line ABCD in the phase diagram of [Figure 10.3](#). The temperature change depends primarily on the pressure drop.

Again, the condensed liquids must be fractionated to meet vapor pressure and composition specifications. This process is most favored when the well-head gas is produced at a very high pressure and can be expanded to sales line pressure with no recompression.

10.4.5 Cryogenic (Turbo-Expander) Plants

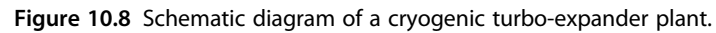
The turbo-expander process for treating gas streams for high liquid recovery was developed in the early 1960s. These plants replace the chiller or J-T

valve with an expansion turbine and are capable of cooling the gas to -160°F (-107°C). The main application of the turbo-expander is to improve the recovery of ethane from natural gas. The process achieves very low temperatures and, therefore, liquefies a substantial portion of the ethane and heavier compounds in natural gas. The various fractions of the liquid stream are recovered by distillation.

The turbo-expander process configurations can vary greatly. They all incorporate various heat exchangers. The gas entering the turbo-expander process must be dehydrated upstream of the plant to a very low water content so that no hydrates form when the low temperatures are reached by the gas being processed. This usually requires a glycol dehydration unit for removing most of the water, followed by a molecular sieve unit to remove virtually all of the water from the feed gas. A common class of molecular sieve used for deep drying has a pore opening of 4 Å. Gas pretreatment can also include CO_2 and H_2S removal.

As the entering gas expands, it supplies work to the turbine shaft, thus reducing the gas enthalpy. A decrease in enthalpy causes a much larger temperature drop than that found in the J-T (constant enthalpy) process. The expansion process is indicated as line “ABC” in the phase diagram in [Figure 10.3](#). The gas is routed through heat exchangers where it is cooled by the residue gas, and condensed liquids are recovered in a cold separation at approximately -90°F (-68°C). These liquids are injected into the de-methanizer at a level where the temperature is approximately -90°F (-68°C). Gas is then expanded (its pressure is decreased from inlet pressure to 225 psig (15.5 barg) through an expansion valve or turbo-expander.

The turbo-expander uses the energy removed from the gas due to the pressure drop to drive a compressor, which helps recompress the gas to sales pressure. The cold gas (-160°F) (-107°C) then enters the de-methanizer column at a pressure and temperature condition where most of the ethane-plus are in the liquid state. As the liquid falls and is heated, the methane is boiled off and the liquid becomes leaner and leaner in methane. Heat is added to the bottom of the tower using the hot discharge residue gas from the compressors to ensure that the bottom liquids have an acceptable Reid vapor pressure (RVP) or methane content. Because of the lower temperatures that are possible, cryogenic plants have the highest liquid recovery levels. [Figure 10.8](#) is an illustration of a simple turbo-expander facility. There are many other arrangements possible, depending on the gas composition and the desired level of liquids recovery. Whether the turbo-expander



is the best choice for removing ethane and heavier HCs from natural gas requires considerable analysis.

Typical Liquid Recovery Levels

- $C_2 = 60\%$
- $C_3 = 90\%$
- $C_4+ = 100\%$

Advantages

- Simple to use
- Easy to package (more expensive than refrigeration)

The design of a turbo-expander unit involves detailed heat and material balances and many flash calculations. Such design calculations are performed by computer simulation programs.



10.5 PROCESS SELECTION

Normally, an economic comparison between viable alternatives will be required; however, the following guidelines are offered.

10.5.1 If NGL Content of the Feed Gas Is Low

- Use the expander process.

10.5.2 If Gases Are Very Rich in NGL

- Simple refrigeration is probably best choice, while expansion is not usually satisfactory.

10.5.3 If Inlet Gas Pressure Is Very High

- Low Temperature Separator (LTS) may be attractive.

10.5.4 If Low Inlet Gas Pressure

- An expander plant or straight refrigeration (if the gas is very rich) is favored.

10.5.5 If Very Low Gas Rates

- Only a very simple process, such as an automatically operated J-T unit, may be justified.

10.5.6 If Large Flow Rates

- A more complex plant with more complex controls and more operating personnel may be justified.

10.5.7 If Remote Wells

- Remote wells may dictate simple operation and processing such as a J-T plant.

A large number of wells may justify a central processing facility with more complex processing.



10.6 FRACTIONATION

Bottoms liquid from any gas plant may be sold as a mixed product. Bottoms liquid may be separated into its various components:

- Ethane
- Propane
- Butane
- Natural gasoline

Fractionation splits liquids into its various components. [Figure 10.9](#) is a simplified flow diagram of a fractionation system. Liquids are cascaded through a

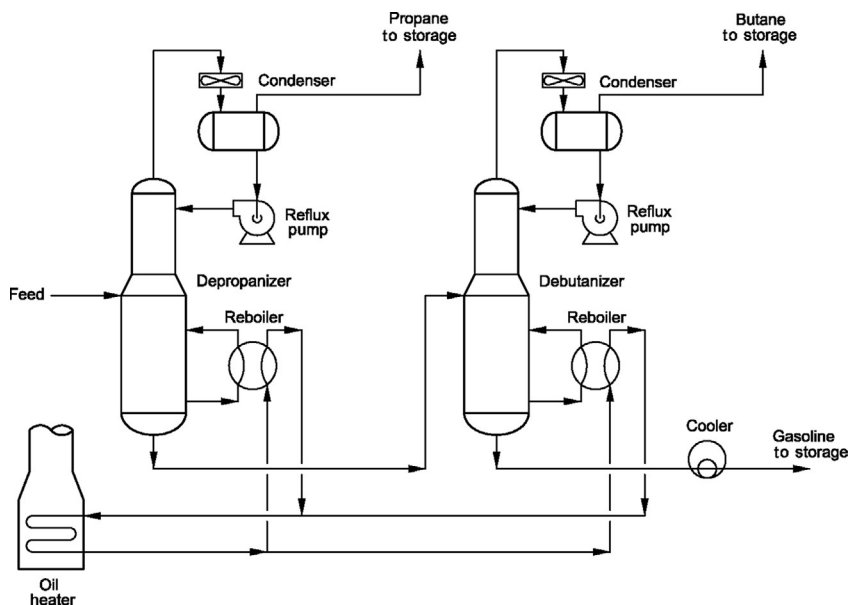


Figure 10.9 Schematic diagram of a fractionation system.

series of distillation towers where successively heavier and heavier components are separated as overhead gas.

Specifications are normally in the form of RVP (controlled by upstream tower) and the amount of heavy ends are set by the fractionator.



10.7 DESIGN CONSIDERATIONS

Proper design requires choosing

- Operating pressure
- Bottoms temperature
- Reflux condenser
- Temperature
- Number of trays

Detail design is normally accomplished by using any one of several commercially available process simulation programs.



Safety Systems

NOMENCLATURE

- d nominal tip diameter, L , in.
 d_i pipe inside diameter, L , in.
 D minimum distance from the midpoint of the flame to the object being considered, L , ft
 D_t tailpipe diameter, L , in the same units as Y
 E fraction of heat radiated
 g acceleration due to gravity, 32.3 ft/s^2
 h height of liquid seal, L , ft
 k ratio of specific heats, C_p/C_v
 K allowable radiation level, BTU/h-ft^2
 L flame length, L , ft
 p maximum allowable header backpressure, m/Lt^2 , psi
 P_{CL} critical pressure at flare tip, m/Lt^2 , psia
 Q heat release (lower heating value), BTU/h
 Q_g gas-flow rate, MMscf/D
 r relative humidity, fraction
 R distance from flare center
 S specific gravity, fraction
 t temperature, T , $^{\circ}\text{F}$
 T temperature, T , $^{\circ}\text{R}$
 \underline{U}_x lateral-wind velocity, L/t
 \underline{U}_j exit gas velocity from stack, L/t
 V gas velocity, L/t , ft/s
 V_S sonic velocity, L/t , ft/s
 W weight flow rate of the vapor/air mixture at distance Y from the end of the tailpipe
 W_f gas-flow rate, lbm/h
 W_o weight flow rate of the relief device discharge, in the same units as W
 x_c horizontal distance from flare tip to flame center, L
 y_c vertical distance from flare tip to flame center, L
 Y distance along the tailpipe axis at which W is calculated, L
 Z gas compressibility at standard conditions, Lt^2/m , psi^{-1}
 Δx horizontal flame distortion caused by lateral wind, L , ft
 Δy vertical flame distortion caused by lateral wind, L , ft
 ΔP_w pressure drop at the tip in inches of water
 ρ sealing-liquid density, lbm/ft^3
 ρ_g density of gas, lbm/ft^3
 τ fraction of heat intensity transmitted



11.1 BASIC SAFETY CONCEPTS

11.1.1 Introduction

Production facilities usually operate according to design. Oil and gas travel from the reservoir to the surface facilities where they are separated, cleaned, and measured and then sent through a pipeline to the end user. During most of this process, everything operates according to plan. Occasionally, problems occur, things break, malfunctions happen, settings change, horns go off, and shut-ins take place. Such problems usually can be solved quickly and easily without negative consequences. Unfortunately, some problems have the potential for serious consequences such as injury to personnel, pollution of the environment, and loss of company assets. Understanding, preventing, or minimizing potential negative consequences requires a fundamental understanding of basic protection concepts and safety analysis.

To develop a safe design, it is necessary to first design and specify all equipment and systems in accordance with applicable codes and standards. Once the system is designed, a process safety shutdown system is specified to ensure that potential hazards that can be detected by measuring process upsets are detected and that appropriate safety actions (normally an automatic shutdown) are initiated. A hazards analysis is then typically undertaken to identify and mitigate potential hazards that could lead to fire, explosion, pollution, or injury to personnel and that cannot be detected as process upsets. Finally, a system of safety management is implemented to ensure the system is operated and maintained in a safe manner by personnel who have received adequate training.

This chapter summarizes the basic protection concepts required for the safe design and operation of a production facility. The chapter begins by developing a hazard tree for a generic production facility and then illustrates how hazards analysis can be used to identify, evaluate, and mitigate process hazards. In addition, this chapter reviews the safety-analysis technique presented in the American Petroleum Institute's (API's) *Recommended Practice (RP) 14C* [1]. While API RP 14C provides guidance on the need for process safety devices, it is desirable to perform a complete hazards analysis of the facility to identify hazards that are not necessarily detected or contained by process safety devices, including those that could lead to loss of containment of hydrocarbons or otherwise lead to fire, explosion, pollution, or injury to personnel. The industry consensus standard, American Petroleum Institute Recommended Practice 14J, *Design and Hazards Analysis for*

Offshore Facilities (API RP 14J), provides guidance as to the use of various hazards-analysis techniques. The chapter describes the management of safety using Safety and Environment Management Programs (SEMP) as defined in API RP 75, *Recommended Practices for Development of a Safety and Environmental Management Program for the Outer Continental Shelf (OCS) Operations and Facilities*, and Safety Case approach as is commonly done in the North Sea. The chapter concludes with a discussion on relief-valve selection and sizing and vent, flare, and relief-systems design.

11.1.2 Basic Protection Concepts

Most threats to safety from production involve the release of hydrocarbons; therefore, the analysis and design of a production-facility safety system should focus on preventing such releases, stopping the flow of hydrocarbons to a leak if it occurs, and minimizing the effects of hydrocarbons should they be released.

11.1.2.1 Prevention

Ideally, hydrocarbon releases should never occur. Every process component is protected with two levels of protection: primary and secondary. The reason for two levels of protection is that if the first level fails to function properly, a secondary level of protection is available.

11.1.2.2 Shut-In

If hydrocarbon releases occur (and, in spite of our best efforts, they sometimes do), inflow to the release site must be shut off as soon as possible. The problem should not be exacerbated with the continued release of additional hydrocarbons. Protective shut-in action is achieved by both the surface safety system (SSS) and the emergency support system (ESS).

11.1.2.3 Minimizing

When hydrocarbons are released, their effects should be minimized as much as possible. This can be accomplished through the use of ignition-prevention measures and ESSs (i.e., the liquid-containment system). If oil spills from a process component, a release of hydrocarbons has occurred. A spill is never good, but component skids and deck drains (if offshore) minimize the effect of a bad situation when the spill would otherwise go into a freshwater stream or offshore waters.

11.1.3 Hazard Tree

A hazard tree identifies potential hazards, determines the conditions necessary for a hazard to exist, determines sources that could create this condition, and breaks the chain leading to the hazard by eliminating the conditions and sources. Because complete elimination is normally not possible, the goal is to reduce the likelihood of occurrence. With statistical analysis, the probability of occurrence can be determined. The effect of a safety procedure or device that reduces the probability of a condition or source occurring also can be quantified with this tool.

A hazard tree is somewhat subjective in that different evaluators may classify conditions and sources differently or they may carry the analysis to further levels of sources. The hazard tree helps the investigator focus attention on all of the aspects to be considered. No matter how the tree is formulated, conclusions reached concerning the design, maintenance, traffic patterns, lighting, and so forth, should be similar.

11.1.4 General-Production-Facility Hazard Tree

Figure 11.1 [2] shows a hazard tree for a generic production facility. It should be equally valid for an offshore or onshore facility. The major hazards are those of oil pollution, fire/explosion, and injury.

11.1.4.1 Oil Pollution

Oil pollution derives from an oil spill that can be caused by one of the conditions shown in Figure 11.1. If an oil spill occurs, pollution could be avoided by installing adequate containment. Requirements for tank dikes, drip pans (offshore), and sumps reduce the probability of oil pollution from most small spills.

One source of an oil spill could be the filling of a vessel that has an outlet to atmosphere until it overflows. Whenever inflow exceeds outflow, the tank can eventually overflow. Another source for a spill is a rupture or sudden inability of a piece of equipment to contain pressure. Events leading to rupture are listed in Figure 11.1. Note that some of these events can be anticipated by sensing changes in process conditions that lead to the rupture. Other events cannot be anticipated from process conditions.

Other sources for oil spills are listed. For example, if a valve is opened and the operator inadvertently forgets to close it, oil may spill out of the system. If there is not a big enough dike around the system, oil pollution will result. It is also possible for oil to spill out the vent/flare system. All pressure vessels

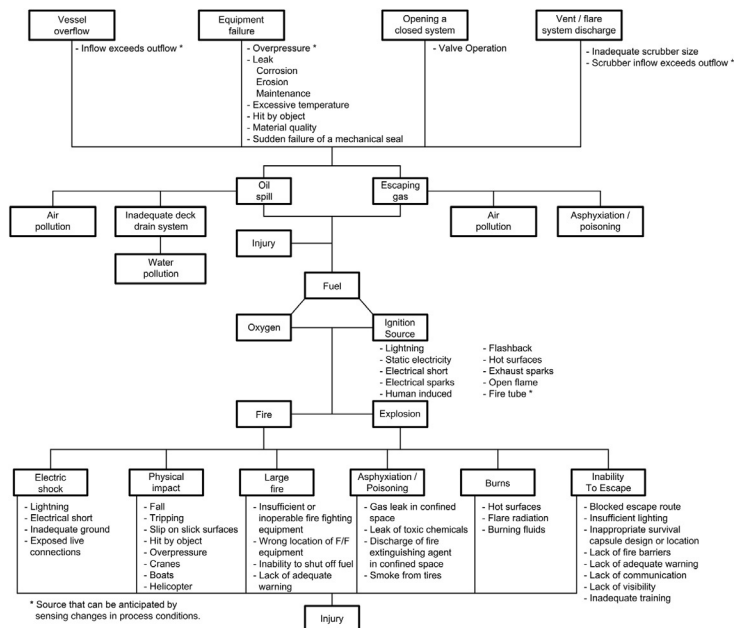


Figure 11.1 Hazard tree for a generic production facility. *Reproduced courtesy of the American Petroleum Institute [2].*

are connected to a relief valve, and the relief valve discharges out a vent or flare system. If the relief scrubber is not adequately sized, or if it does not have a big enough dump rate, oil will go out the vent system.

11.1.4.2 Fire/Explosion

Fire and explosion are much more serious events than pollution. For one thing, fire and explosion can create catastrophes that will lead to pollution anyway, but for another thing, they can injure people. We clearly want to have more levels of safety (that is, a lower probability of occurrence) in the chain leading to fire or explosion than is necessary in the chain leading to pollution. That is, whatever the acceptable risk for oil pollution, a lower risk is required for fire or explosion.

For fire or explosion to occur, fuel, and an ignition source, oxygen, and time to mix them all together are needed. An oil spill or gas leak can provide fuel for a fire/explosion. If any of these elements can be eliminated with 100% assurance, the chain leading to fire or explosion will be broken. For example, if oxygen can be kept out of the facility, then there can be no fire or explosion. Eliminating oxygen can be done inside the equipment by designing a gas

blanket and ensuring positive pressure. For practical purposes it cannot be done outside the equipment, as a human interface with the equipment is desired.

Fuel cannot be completely eliminated, though the inventory of combustible fuels can be kept to a minimum. Oil and gas will be present in any production facility, and either an oil spill or escaping gas can provide the fuel needed. Escaping gas can result from rupture, opening a closed system, or gas that is normally vented. The amount of fuel present can be minimized by preventing oil spills and gas leaks.

Ignition sources are numerous, but it is possible to minimize them. Lightning and static electricity are common ignition sources in production facilities, especially tank vents. It is not possible to anticipate the ignition by sensing changes in process conditions, but gas blankets, pressure vacuum valves, and flame arrestors can be installed to ensure that flame will not flash back into the tank and create an explosion. Electrical shorts and sparks are also sources of ignition. These are kept isolated from any fuel by a series of rules and regulations for the design of electrical systems. In the United States, the National Electrical Code and the API Recommended Practices for Electrical Systems are used to minimize the danger of these ignition sources. Human-induced ignition sources include welding and cutting operations, smoking, and hammering (which causes static electricity). Flash back is also a source of ignition. In some vessels, a flame exists inside a fire tube. If a fuel source develops around the air intake for the fire tube, the flame can propagate outside the fire tube and out into the open. The flame would then become a source of ignition for any more fuel present and could lead to a fire or explosion. This is why flame arrestors are not required on natural draft fire tubes.

Hot surfaces are another common source of ignition. Engine exhaust, turbine exhaust, and engine manifold on engine-driven compressors may be sufficiently hot to ignite oil or gas. A hot engine manifold can become a source of ignition for an oil leak. An engine exhaust can become a source of ignition for a gas escape.

Exhaust sparks from engines and burners can be a source of ignition. Any open flame in the facility can also be a source of ignition.

Fire tubes, especially in heater treaters, where they can be immersed in crude oil, can become a source of ignition if the tube develops a leak, allowing crude oil to come in direct contact with the flame. Fire tubes can also be a source of ignition if the burner controls fail and the tube overheats or if the pilot is out and the burner turns on when there is a combustible mixture in the tubes.

Because these ignition sources cannot be anticipated by sensing changes in process conditions, and since oxygen is always present, a hazards analysis must concentrate on reducing the risk of oil spill and gas leak when any of these ignition sources are present. The hazards analysis must focus on reducing the probability that the ignition source will exist at the same location as an oil spill or gas leak.

11.1.4.3 Injury

Injury can occur directly from an explosion, an out-of-control fire, or one of the other conditions shown in [Figure 11.1](#). A fire can lead directly to injury, but normally there needs to be several contributing events before the fire becomes large enough to lead to injury. For example, if a fire develops and there is sufficient warning, there should be sufficient time to escape before injury results. If there is sufficient warning before a fire develops, there should be enough time to escape before injury occurs. If the fuel can be shut off and adequate fire-fighting equipment is present to control the fire before it becomes a large fire, the probability of injury is minimal.

When an explosion occurs, however, it can directly cause injury. A substantial cloud of gas can accumulate before the combustible limit reaches an ignition source. The force of the explosion as the cloud ignites can be substantial.

There are other ways to injure people, such as physical impact due to falling, tripping, slipping on a slick surface, or being hit by an object or by direct physical impact from a rupture. Asphyxiation can occur, especially when dealing with toxic chemicals.

Electric shock and burns can also lead to injury. Burns can occur by touching hot surfaces. They can also occur from radiation.

The inability to escape increases the probability of injury from any of these conditions. All the conditions are more likely to lead to injury the longer personnel are exposed to the situation; therefore, escape routes, lighting, appropriate survival capsules/boats (if offshore), and fire barriers all lead to a reduction in the probability of injury.

11.1.5 Severity of Source

The hazard tree helps identify the severity of a source that can lead to a hazardous condition. Some of these sources are discussed here.

11.1.5.1 Overpressure

Overpressure can lead directly to all three hazards. It can lead directly and immediately to injury; it can lead to fire/explosion if there is an ignition source; and it can lead to pollution if there is insufficient containment. Because of the hazard potential, a very good level of assurance is needed that the probability of overpressure occurring is very small.

11.1.5.2 Fire Tubes

Fire tubes can lead to fire/explosion if there is a leak of crude oil or glycol into the tubes or if there is a failure of the burner controls. An explosion may be sudden and lead directly to injury; therefore, a high degree of safety is required.

11.1.5.3 Excess Temperature

Excess temperature can cause premature equipment failure at a pressure below its maximum design working pressure. Excess temperature can create a leak, potentially leading to fire or explosion if gas leaks or to oil pollution if oil leaks. This type of failure should be gradual, giving off a warning as it develops, and thus does not require as high a degree of protection as those mentioned previously.

11.1.5.4 Leaks

Leaks rarely lead directly to personnel injury, but they can lead to fire/explosion if there is an ignition source and to oil pollution if there is inadequate containment. The immediacy and magnitude of the developing hazard will be less than with overpressure; thus, although it is necessary to protect against leaks, this protection will not require the same level of safety required for overpressure.

11.1.5.5 Inflow Exceeds Outflow

Inflow exceeding outflow can lead to oil pollution if there is inadequate containment and can lead to fire/explosion and, thus, to injury if an oil spill occurs. This condition is more time dependent and lower in magnitude of damage; therefore, an even lower level of safety is acceptable.

11.1.6 Need for Other Protection Devices

The hazard tree also helps identify other protection devices to include in equipment design that may minimize the possibility that a source will

develop into a hazardous condition. Additional protection devices that might be included are flame arrestors, stack arrestors, gas detectors, fire detectors, and manual shutdown stations. A hazards analysis can determine the need for safety devices and safety systems.



11.2 DEVELOPING A SAFE PROCESS

In going through this hazard tree, it can be seen that many of the sources and conditions leading to the three major hazards have nothing to do with the way in which the process is designed. Many sources cannot be anticipated by sensing a condition in the process. For example, it is not possible to put a sensor on a separator that keeps someone who is approaching the separator to perform maintenance from falling. Another way of stating this is that many of the sources and conditions identified on the hazard tree require design considerations that do not appear on mechanical flow diagrams. The need for proper design of walkways, escape paths, electrical systems, fire-fighting systems, insulation on piping, and so forth, is evident on the hazard tree. In terms of developing a process safety system, only those items that are starred in the hazard tree can be detected and therefore defended against.

This point must be emphasized because it follows that a production facility that is designed with a process shut-in system as described in API RP14C is not necessarily “safe.” It has an appropriate level of devices and redundancy to reduce the sources and conditions that can be anticipated by sensing changes in process conditions. However, much more is required from the design of the facility if the overall probability of any one chain leading to a hazard is to be acceptable. That is, API RP14C is merely a document that has to do with safety analysis of the process components in the production facility. It does not address all the other concerns that are necessary for a “safe” design.

The starred items in the hazard tree are changes in process conditions that could develop into sources and lead to hazards. These items are identified in [Table 11.1](#) in the order of their severity.

Overpressure can lead directly to all three hazards. It can lead directly and immediately to injury, to fire or explosion if there is an ignition source, and to pollution if there is not enough containment. Therefore, we must have a very high level of assurance that overpressure is going to have a very low frequency of occurrence.

Table 11.1 Sources Associated with Process System Changes

Source	Hazards	Contributing Source of Condition
Overpressure	Injury	None
	Fire/explosion	Ignition source
	Pollution	Inadequate containment
Leak	Fire/explosion	Ignition source
	Oil pollution	Inadequate containment
Fire tubes	Fire/explosion	Fuel
Inflow exceeds outflow	Oil pollution	Inadequate containment
Excessive temperature	Fire/explosion	Ignition source
	oil pollution	Inadequate containment

Fire tubes can lead to fire or explosion if there is a leak of crude oil into the tubes or failure of the burner controls. An explosion could be sudden and lead directly to injury. Therefore, a high level of safety is required.

Excessive temperature can lead to premature failure of an item of equipment at pressures below its design maximum working pressure. Such a failure can create a leak, potentially leading to fire or explosion if gas is leaked or to oil pollution if oil is leaked. This type of failure should be gradual, with warning as it develops, and thus does not require as high a degree of protection as those previously mentioned.

Leaks cannot lead directly to personal injury. They can lead to fire or explosion if there is an ignition source and to oil pollution if there is inadequate containment. Both the immediacy of the hazard developing and the magnitude of the hazard will be smaller with leaks than with overpressure. Thus, although it is necessary to protect against leaks, this protection will not require the same level of safety that is required to protect against overpressure.

Inflow exceeding outflow can lead to oil pollution if there is inadequate containment. It can lead to fire or explosion and thus to injury by way of creating an oil spill. This type of accident is more time-dependent and lower in magnitude of damage, and thus an even lower level of safety will be acceptable.

The hazard tree also helps identify protection devices to include in equipment design that may minimize the possibility that a source will develop into a condition. Examples would be flame arrestors and stack arresters on fire tubes to prevent flash back and exhaust sparks, gas detectors to sense the presence of a fuel in a confined space, and fire detectors and manual shutdown stations to provide adequate warning and to keep a small fire from developing into a large fire.

11.2.1 Primary Defense

Before proceeding to a discussion of the safety devices required for the process, it is important to point out that the primary defense against hazards in a process system design is the use of proper material of sufficient strength and thickness to withstand normal operating pressures. This is done by designing the equipment and piping in accordance with accepted industry design codes. If this is not done, no sensors will be sufficient to protect from overpressure, leak, and so on. For example, a pressure vessel is specified for 1480 psi maximum working pressure, and its relief valve will be set at 1480 psi. If it is not properly designed and inspected, it may rupture before reaching 1480 psi pressure. The primary defense to keep this from happening is to use the proper codes and design procedures and to ensure that the manufacture of the equipment and its fabrication into systems are adequately inspected. In the United States, pressure vessels are constructed in accordance with the ASME Boiler and Pressure Vessel Code, and piping systems are constructed in accordance with one of the ASME Piping Codes discussed in Volume 3.

It is also important to ensure that corrosion, erosion, or other damage has not affected the system to the point that it can no longer safely contain the design pressure. Maintaining mechanical integrity once the system has been placed in service is essential to maintaining a safe facility.

11.2.2 Failure Mode Effect Analysis

One of the procedures used to determine which sensors are needed to sense process conditions and protect the process is called a failure mode effect analysis (FMEA). Every device in the process is checked for its various modes of failure. A search is then made to ensure that there is a redundancy that keeps an identified source or condition from developing for each potential failure mode. The degree of required redundancy depends on the severity of the source as previously described. [Table 11.2](#) lists failure modes for various devices commonly used in production facilities.

In applying FMEA, a mechanical flow diagram must first be developed. As an example, consider the check valve on a liquid dump line. It can fail one of three ways—it can fail to close, it can leak internally, or it can leak externally. The FMEA will investigate the effects that could occur if this particular check valve fails to close. Assuming this happens, some redundancy that keeps a source from developing must be located in the system. Next, the process would be evaluated for the second failure mode, that is, what occurs

Table 11.2 Failure Modes of Various Devices

Sensors		Signal/Indicator	
FTS	Fail to see	FTI	Fail to indicate
OP	Operate prematurely		
Check Valves		Switch	
FTC	Fail to close (check)	FS	Fail to switch
Lin	Leak internally	FC	Fail close
Lex	Leak externally	FO	Fail open
Orifice Plates (Flow Restrictor)		Engine	
FTR	Fail to restrict	FTD	Fail to deliver
BL	Block	FXP	Deliver excess power
Pumps		Transformer	
FTP	Fail to pump	FTF	Fail to function
POP	Pump to overpressure		
LEX	Leak externally		
Controllers		General	
FTCL	Fail to control level	OF	Overflow
FTCT	Fail to control temperature	NP	Not processed
FTCF	Fail to control flow	NS	No signal
OP	Operate prematurely	FP	Fail to power
FTCLL	Fail to control low level	MOR	Manual override
FTCHL	Fail to control high level	NA	Not applicable
FTRP	Fail to reduce pressure		
FTCP	Fail to control pressure	Rupture Disc	
FTAA	Fail to activate alarms	RP	Rupture prematurely
Valves		FTO	Fail to open
FO	Fail open	LEX	Leak externally
FC	Fail close		
FTO	Fail to open	Meter	
FTC	Fail to close	FTOP	Fail to operate Properly
Lin	Leak internally	LEX	Leak externally
Lex	Leak externally	BL	Block
		Timer	
		FTAP	Fail to activate pump
		FTSP	Fail to stop pump

if the check valve leaks internally. Next, the process would be evaluated for the third failure mode of this check valve. Check valves are easy. A controller has nine failure modes, and a valve has six.

In order to perform a complete, formal FMEA of a production facility, each failure mode of each device must be evaluated. A percentage failure rate and cost of failure for each mode for each device must be calculated. If the

risk discounted cost of failure is calculated to be acceptable, then there are the proper numbers of redundancies. If that cost is not acceptable, then other redundancies must be added until an acceptable cost is attained.

It is obvious that such an approach would be lengthy and require many pages of documentation that would be difficult to check. It is also obvious that such an approach is still subjective in that the evaluator must make decisions as to the consequences of each failure, the expected failure rate, and the acceptable level of risk for the supposed failure.

This approach has been performed on several offshore production facilities with inconsistent results. That is, items that were identified by one set of evaluators as required for protection in one design were not required by another set of evaluators in a completely similar design. In addition, potential failure of some safety devices on one facility caused evaluators to require additional safety devices as back-up, while the same group in evaluating a similar installation that did not have the initial safety devices at all did not identify the absence of the primary safety device as a hazard or require back-up safety devices.

It should be clear that a complete FMEA approach is not practical for the evaluation of production facility safety systems. This is because (1) the cost of failure is not as great as for nuclear power plants or rockets, for which this technology has proven useful; (2) production facility design projects cannot support the engineering cost and lead time associated with such analysis; (3) regulatory bodies are not staffed to be able to critically analyze the output of an FMEA for errors in subjective judgment; and most importantly, (4) there are similarities to the design of all production facilities that have allowed the industry to develop a modified FMEA approach that can satisfy all these objections.

11.2.3 Modified FMEA Approach

The modified FMEA approach evaluates each piece of equipment (not each device) as an independent unit, assuming worst-case conditions of input and output. Separators, flowlines, heaters, compressors, and other equipment function in the same manner no matter the specific design of the facility. That is, they have level, pressure and temperature controls, and valves. These are subject to failure modes that impact the piece of equipment in the same manner. Thus, if an FMEA analysis is performed on the item of equipment standing alone, the FMEA will be valid for that component in any process configuration.

Furthermore, once every process component has been analyzed separately for worst case, stand-alone conditions, there is no additional safety risk created by joining the components into a system. That is, if every process component is fully protected based on its FMEA analysis, a system made up of several of these components will also be fully protected.

It is even possible that the system configuration is such that protection furnished by devices on one process component can protect others. That is, devices that may be required to provide adequate protection for a component standing alone may be redundant once all components are assembled in a system. This procedure is outlined below:

1. For each piece of equipment (process component), develop an FMEA by assuming in turn that each process upset that could become a potential source occurs. That is, assume a control failure, leak, or other event leading to a process upset.
2. Provide a sensor that detects the upset and shuts-in the process before an identified source of condition develops. For example, if the pressure controller fails and the pressure increases, provide a high-pressure sensor to shut-in the process. If there is a leak and the pressure decreases, provide a low-pressure sensor to shut-in the process.
3. Apply FMEA techniques to provide an independent back-up to the sensor as a second level of defense before an identified hazard is created. The degree of reliability of the back-up device will be dependent upon the severity of the problem. For example, since overpressure is a condition that can lead to severe hazards, the back-up device should be extremely reliable. Typically, a high-pressure sensor would be backed up by a relief valve. In this case, a relief valve is actually more reliable than the high-pressure sensor, but it has other detriments associated with it. Oil leakage, on the other hand, is not as severe. In this case, a drip pan to protect against oil pollution may be adequate back-up.
4. Assume that two levels of protection are adequate. Experience in applying FMEA analysis to production equipment indicates that in many cases only one level of protection would be required, given the degree of reliability of shutdown systems and the consequences of failure. However, it is more costly in engineering time to document that only one level is required for a specific installation than it is to install and maintain two levels. Therefore, two levels are always specified.
5. Assemble the components into the process system and apply FMEA techniques to determine if protection devices on some components provide redundant protection to other components. For example, if there

Table 11.3 FMEA of a Separator

Undesirable Event	Primary	Secondary
Overpressure	PSH	PSV
Large gas leak	PSL and FSV	ASH, minimize ignition sources
Large oil leak	LSL and FSV	Sump tank (LSH)
Small gas leak	ASH, Minimize Ignition Source	Fire detection
Small oil leak	Sump Tank (LSH)	Manual observation
Inflow exceeds outflow	LSH	Vent scrubber (LSH)
High temperature	TSH	Leak detection devices

are two separators in series, and they are both designed for the same pressure, the devices protecting one from overpressure will also protect the other. Therefore, there may be no need for two sets of high-pressure sensors.

The application of this procedure is best seen by performing an FMEA on a simple two-phase separator. [Table 11.3](#) lists those process upsets that can be sensed before an undesirable event leading to a source of condition occurs. For overpressure, primary protection is provided by a high-pressure sensor that shuts in the inlet (PSH). If this device fails, secondary protection is provided by a relief valve (PSV).

A large leak of gas is detected by a low-pressure sensor (PSL) that shuts in the inlet, and a check valve (FSV) keeps gas from downstream components from flowing backward to the leak. Similarly, a large oil leak is detected by a low-level sensor (LSL) and a check valve. Back-up protection is provided by a sump tank and its level safety high (LSH) for an oil leak. That is, before an oil spill becomes pollution there must be a failure of a second sensor. Back-up protection for a gas leak then becomes the fire detection and protection equipment if the small leak were to cause a fire. There is no automatic back-up to the sump tank LSH for a small oil leak. Manual intervention, before containment is exceeded and oil pollution results, becomes the back-up.

The primary protection for high temperature, which could lower the maximum allowable working pressure (MAWP) below the pressure safety valve (PSV) setting, is a high-temperature sensor (TSH), which shuts in the inlet or the source of heat. Back-up protection is provided by leak detection devices.

Inflow exceeding outflow is sensed by a level safety high (LSH). Back-up protection is furnished by the PSH (to keep the relief valve from operating) or an LSH in a downstream vent scrubber if the vessel gas outlet goes to

atmosphere. That is, a vent scrubber must be installed downstream of any vessel that discharges directly to atmosphere.

Once the FMEA is completed, the specific system is analyzed to determine if all the devices are indeed needed. For example, if it is not possible for the process to overpressure the vessel, these devices are not required. If it is impossible to heat the vessel to a high enough level to effect its maximum working pressure, the TSH can be eliminated.



11.3 API RP 14C

11.3.1 Overview

API RP 14C [1] is a safety-analysis approach based on a number of traditional hazards-analysis techniques such as failure mode effects analysis (FMEA) and hazard-and-operability studies. The purpose of a safety analysis is to identify undesirable events that might pose a threat to safety, and define reliable protection measures that will prevent such events or minimize their effects should they occur. Potential threats to safety are identified through proven hazards-analysis techniques that have been adapted to hydrocarbon-production processes. Recommended protective measures are common industry practices proved through many years of operating experience. The hazards analysis and protective measures have been combined into a “safety analysis” for onshore and offshore production facilities.

The API RP 14C [1] safety analysis is based on the following premises.

- Process components function in the same manner regardless of specific facility design.
- Each process component is analyzed for “worst case” input and output conditions.
- If fully protected when analyzed standing alone, the analysis will be valid for that component in any configuration.
- If every component is protected, the system will be protected.
- When components are assembled into a system, some devices can be eliminated.

The major benefits of this analysis are concise, easy-to-audit documentation; minimized subjective decisions; and consistent results. The remainder of [Section 11.3](#) explains the basic concepts of protection used in the analysis, discusses the methods of analyzing the process, and establishes design criteria for an integrated safety system. The entire production process is covered, and a step-by-step summary for performing a safety analysis is provided.

11.3.2 Process Variables

There are four main process variables in upstream production facilities: pressure, liquid level, temperature, and flow. A variable fluctuates between a lower and an upper extreme value. For example, the liquid level within a vessel can fluctuate from the bottom of the vessel (empty) to the top (full). Process variables allow movement of the fluids through the process components while simultaneously achieving the degree of separation required for sales or water disposal.

11.3.3 Process Components

A process component is any piece of equipment that handles hydrocarbons. Identifying all the components that handle hydrocarbons in a production facility would be overwhelming. Instead of listing components by their common name, *RP 14C* [1] lists components by their functions, thus decreasing the number of names from hundreds to only ten. Regardless of what a piece of equipment is called, it can be described as one of the following ten process components: wellheads and flowlines, wellhead injection lines, headers, pressure vessels, atmospheric vessels, fired and exhaust-heated components, pumps, compressors, pipeline, and shell-and-tube heat exchangers.

11.3.4 Normal Operating Ranges

Whenever hydrocarbons are present in a process component, each of the four main process variables take on some value. Values at which the variables can be found when things are going smoothly are called normal values. For example, the pressure on a flowline will fluctuate from reading to reading within a specified period of time (e.g., for a 1-h period the readings may be 950 psi at 1300 h, 1010 psi at 1340 h, and 979 psi at 1400 h). As long as flow is occurring, the liquid level within a process component is changing. For example, in a separator's oil bucket, the level will steadily rise until the dump valve opens and drains some of the oil, at which time the liquid level falls until the dump valve closes. Within each process component, each variable has a normal operating range instead of a single normal value.

One of the cornerstones of facility protection lies in protecting each component against certain undesirable events that are closely related to the four main process variables. For example, if the pressure within a

component were to become too high, a component could rupture; a pressure too low within a component could indicate a leak. A liquid level within a component that is too high or too low could cause problems as well as indicate equipment failure.

Production operators establish normal operating ranges. The principal concern with the four main process variables is that their sensing devices, which respond to conditions outside normal ranges, have enough time to respond before problems occur. For example, the normal range for a separator's liquid level can be wherever the operator wants it, provided that the level safety high (LSH) can shut off inflow before liquid overflow occurs and that the level safety low can respond before the level has completely disappeared and allows gas to flow out of the liquid outlet (gas blowby). The normal operating range for a component's pressure can be established by attaching a pressure recorder to the component and recording the pressure variations over time.

Maintaining normal operating ranges requires normal process flow. When the four main process variables are kept within their normal ranges, process flow is occurring. Process flow is maintained by chokes, regulators, controllers, and the influence of the main process variables on each other; therefore, normal operating ranges are maintained by the same things.

11.3.5 Abnormal Operating Conditions

On average, process variables are found within their normal operating ranges, but horns do go off and shut-ins do occur. When chokes and controllers that normally keep the process variables in their normal ranges fail to function properly, the process variable being controlled can be outside its normal operating limits. Whenever a process variable exceeds its normal range, it is said to be in an abnormal condition. For example, in a component with a normal operating range of 800-900 psig, pressure >900 psig or <800 psig are abnormal conditions. A liquid level above or below the point at which the dump valve opens or closes is an abnormal condition.

What is the significance of an abnormal condition? In reality, when the normal operating range is exceeded by only a small amount, it makes very little difference to the operation of the facility. However, if a variable exceeds its normal operating range at all, it could continue to escalate with potentially disastrous results. Operators are concerned mainly about the consequences that might result if abnormal conditions become extreme.

11.3.6 Consequences

Several consequences can result from abnormal operating conditions. At best, there will be only a horn and a shut-in. The most serious consequences are injury to personnel, pollution, and loss of company assets. Abnormal conditions do not always develop into a serious consequence, but it could happen. According to API *RP 14C* [1], serious consequences usually are preceded by some abnormal condition. Abnormal conditions that are not dealt with quickly can escalate into worst-case scenarios.

11.3.7 Causes

The major causes of abnormal conditions are equipment failure or malfunction and human error. Examples of equipment failure or malfunction are chokes that become enlarged through contact with excessive sand in the flow stream, dump valves that hang open or stay closed, and regulators or controllers that change adjustment because of vibration. Human error can occur if an operator repairing a dump valve does not want to shut in to finish the job and uses the dump valve's bypass line. If the operator fails to monitor the liquid level properly while the bypass valve is open, the liquid level in the component could get too high or too low. Human error also can occur if the operator monitored the level accurately but forgot to check to see if the newly repaired dump valve was operating properly.

11.3.8 Prevention

The actual causes of abnormal conditions are varied and numerous. API *RP 14C* [1] provides an analysis technique to identify potential abnormal conditions and prevent them from occurring.

11.3.9 Effects of Hydrocarbon Releases

Abnormal operating conditions could result in injury to personnel, pollution, and loss of assets. Whenever any of these worst-case consequences is at its most serious, the release of hydrocarbons is usually involved. While pollution of any type is undesirable, hydrocarbon pollution is the most serious. The May 1989 *Exxon Valdez* incident is a prime example of the attention drawn to and the expenses involved with hydrocarbon pollution in navigable waterways. Often, injury to personnel on a major scale also involves the release of hydrocarbons. Hydrocarbon releases alone are often sufficient to cause injury to personnel (e.g., whenever H_2S is involved).

Worst of all is a fire caused by or fed by hydrocarbon releases. An explosion or fire can cause extensive damage to equipment and personnel, which can result in extensive injury, pollution, and facility damage. Offshore platforms have melted to the water line because of released hydrocarbons, as occurred in the Piper Alpha incident in the North Sea in the summer of 1988. Onshore facilities have been completely leveled to the ground because of released hydrocarbons, as occurred in the Phillips incident in Pasadena, Texas, in 1988.

11.3.10 Safety Devices

Safety devices offer a solution for hydrocarbon releases. Specific devices have been developed to protect production facilities. As these devices became more common, industry standards, such as names, symbols and identification, and installation locations, were established. API *RP 14C Sec. 2* [1] summarizes surface-production-facility-related standards.

11.3.10.1 Names

Before installing a specific safety device, a standard way of referring to it is needed. API *RP 14C* presents two groups of safety devices: “common” (i.e., typical oilfield) names such as check valve or pop-off valve and “proper” names from the Instrument Society of America (ISA), such as flow safety valve or PSV. With few exceptions, every ISA name includes the measured or initiating variable as the first part of the name and the word safety as the second part of the name. The third and usually final part of the name refers to either the device itself (i.e., valve or element) or to the type of function the device performs (i.e., high or low).

ISA device names usually are abbreviated with the first letter of each part of the name. If a single component has two or more of the same kind of device on it, each device is differentiated by the addition of a number or letter following the device’s letters (LSH1 and LSH2, for example). The same convention is used for all safety devices. See [Figure 11.2](#) [1].

11.3.10.2 Symbols

Process flow diagrams must show safety devices. A graphic symbol represents each safety device. These symbols save space on the diagram and make the appearance neater. [Figure 11.3](#) [1] contains standardized symbols used in hydrocarbon-facility diagrams.












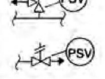
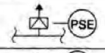
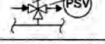
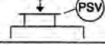

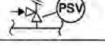
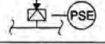




SENSING AND SELF-ACTING DEVICE				
VARIABLE	SAFETY-DEVICE DESIGNATION		SYMBOL	
	COMMON	INSTRUMENT SOCIETY OF AMERICA (ISA)	SINGLE DEVICE	COMBINATION DEVICE
Backflow	Check valve	Flow Safety Valve		
Burner flame	Burner-flame detector	Burner Safety Low		
Flow	High-flow sensor	Flow Safety High		
	Low-flow sensor	Flow Safety Low		
Level	High-level sensor	Level Safety High		
	Low-level sensor	Level Safety Low		
Pressure	High-pressure sensor	Pressure Safety High		
	Low-pressure sensor	Pressure Safety Low		
	Pressure-relief or safety valve	Pressure Safety Valve		
	Rupture disk or safety head	Pressure Safety Element		
Pressure or vacuum	Pressure/vacuum-relief valve	Pressure Safety Valve		
	Pressure/vacuum-relief manhole cover	Pressure Safety Valve		
	Vent	None		
Vacuum	Vacuum-relief valve	Pressure Safety Valve		
	Rupture disk or safety head	Pressure Safety Element		
Temperature	High-temperature sensor	Temperature Safety High		
	Low-temperature sensor	Temperature Safety Low		
Flame	Flame or stack arrestor	None		

Figure 11.2 Safety-device identification examples. *Reproduced courtesy of the American Petroleum Institute [1].*

11.3.11 Production-Process Safety Systems

Production-process safety systems provide a more extensive level of protection than an individual device. They include end devices and auxiliary devices, which are important not only to the system itself but also to the safety of the facility. A brief overview of these systems follows.

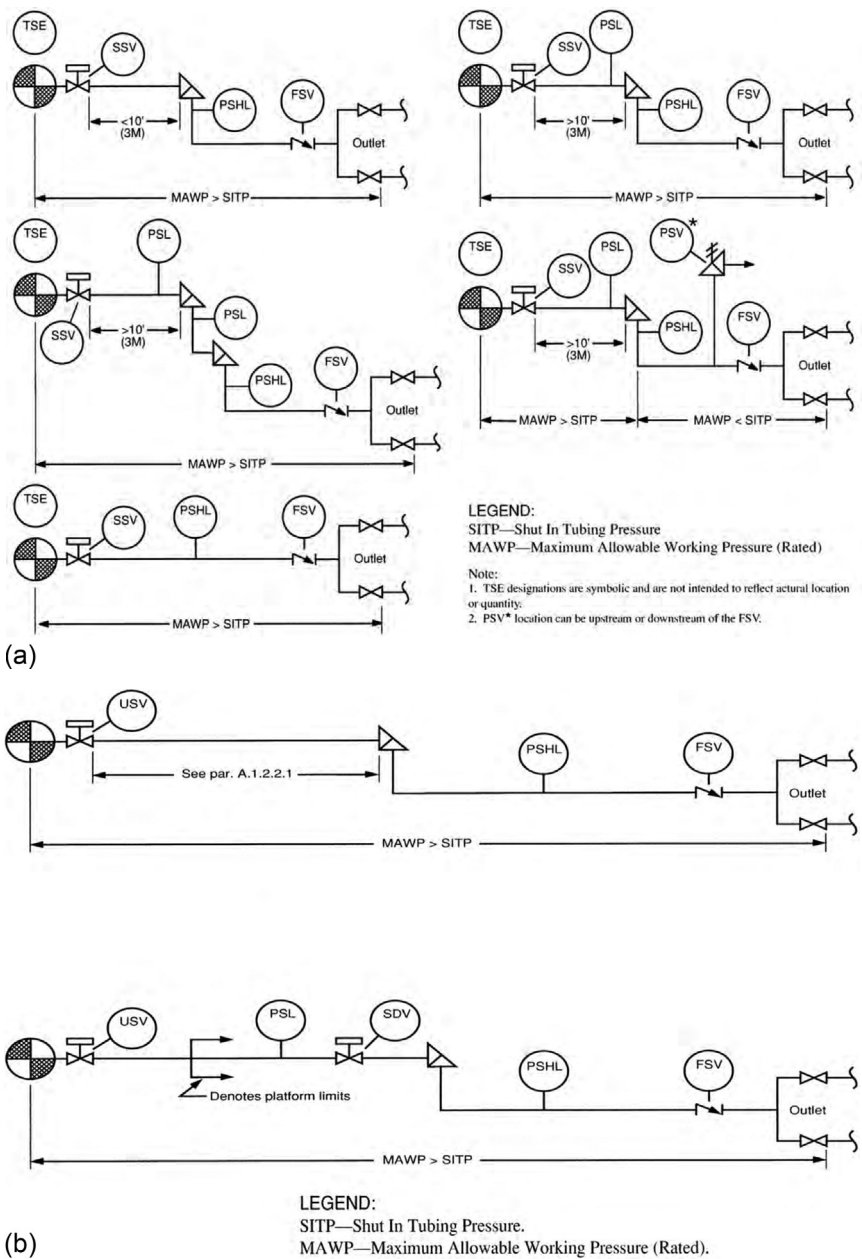


Figure 11.3 Safety-device tables. *Reproduced courtesy of the American Petroleum Institute [1].*

11.3.11.1 Surface Safety System

The SSS consists mostly, but not exclusively, of sensing-type individual safety devices. Devices respond to one of the four major variables. The main purpose of the SSS is to prevent the initial release of hydrocarbons and to shut in additional flow of the hydrocarbons already released. The SSS consists of three major components: sensing devices, relay devices, and end devices. Some devices both sense and respond as an end device (check valves, relief valves, and so forth).

The SSS incorporates various sensing devices. When an abnormal condition is detected, the sensing device sends a signal to an end device. The end device diverts or shuts off flow, sounds an alarm, or takes some other corrective action. For example, if a component's dump valve freezes in the closed position, the liquid level within the component will rise. When it rises high enough, the component's LSH will sense the high level and send a signal that shuts in the wells flowing into the component. The same signal usually will also sound an alarm to notify facility personnel of the shut-in.

11.3.11.2 Emergency Support System

The ESS consists of seven major subsystems, all of which help protect the facility and environment. The main purpose of ESSs is to shut in additional flow and minimize the effects of hydrocarbons that have already been released. The API realizes that hydrocarbon releases ideally would be prevented through the use of sensing devices (i.e., the SSS), but the API also knows that there will be times in which hydrocarbons are released in spite of the SSS. To address this problem, the API mandates a backup means of protecting the facility. The ESS is a major part of those backup efforts. The seven subsystems that make up the ESS are an emergency-shutdown (ESD) system, a fire-detection system, a combustible-gas detection system, adequate ventilation, a liquid containment system, sumps, and subsurface safety valves.

11.3.11.3 Other Support Systems

Two additional systems are required to make a facility as safe as possible. They are the pneumatic supply system and a system for discharging gas to the atmosphere (blowdown/vent). The pneumatic supply system provides the power to operate most of the other systems. The blowdown/vent system provides a means for directing unwanted gas away from the facility while capturing as many liquid hydrocarbons as possible and thereby reducing pollution levels.

11.3.12 Ignition-Prevention Measures

Ignition-prevention measures are designed to prevent released hydrocarbons from being ignited, thereby minimizing the effects of released hydrocarbons. They accomplish this task through four approaches: ventilation, compliance with all applicable electrical codes, locating equipment in areas where exposure to inadvertently released hydrocarbons is minimized, and hot surface protection. Refer to *RP 14C*, paragraph 4.2.4, [1] for more information about these measures.

11.3.13 Undesirable Events

Abnormal operating conditions can lead to one or more undesirable events that, in turn, could lead to injury to personnel, pollution of the environment, and damage to the facility or its equipment. Safety devices and safety systems are added to prevent undesirable events and they provide the last chance to prevent worst-case consequences from occurring. At each stage, action can be taken to keep the main process variables from resulting in worst-case consequences. Chokes and controllers keep the variables within their normal ranges. Once the variables exceed their normal ranges, safety devices respond to keep the variables from getting further out of range. If the undesirable-event stage is reached, there are still ways of preventing or lessening the chance of the occurrence of worst-case consequences (e.g., ESS). Even though they occur less frequently than either normal or abnormal conditions, undesirable events are much more likely to lead to worst-case consequences than either of the other two conditions.

Eight undesirable events can be identified by looking at all the possible ways injury, pollution, and loss of company assets could occur. The process is similar to that used to identify the ten process components. Each of the eight undesirable events can be examined further to determine the most common causes of the undesirable event, the effects of the undesirable event, detectable abnormal conditions that usually precede the undesirable event, the most effective primary and secondary protective devices that could prevent the undesirable event, and the optimal location for the placement of the required safety device.

By studying each of these undesirable events, information can be gained to make a facility safe. For example, by knowing the possible causes of a particular undesirable event, those possible causes can be monitored and often corrected before they develop into an undesirable event. Knowing about the

possible effects of each undesirable event allows for a more rapid or more appropriate response to the undesirable event. Information about the detectable abnormal condition provides a tool for better monitoring and provides information about which types of safety devices can be used to warn of an impending undesirable event. Primary and secondary protection information assists in determining which safety devices are best for that particular undesirable event. Location data provide information on where the safety devices must be positioned for the most effective protection.

API RP 14C [1] does an excellent job of describing this information. It starts by defining an undesirable event as “an adverse occurrence in a process component which poses a threat to safety.” There can be many different types of “threat(s) to safety.” These can range from minor to the catastrophic. API defines undesirable events with catastrophic threats in mind.

The eight undesirable events identified by *RP API 14C* [1] are overpressure, leak, liquid overflow, gas blowby, underpressure, excess temperature (fire and exhaust-heated components), direct ignition source, and excess combustible vapors in the firing chamber (fired components). The following issues are key points about undesirable events.

Worst-case threats to safety originating in process components are usually preceded by one or more of the eight undesirable events.

Each undesirable event has a cause that is usually, but not always, preceded by an abnormal condition. The abnormal condition, in turn, is usually detectable.

Primary protection must be provided to either prevent the undesirable event from occurring or to minimize the effects of the undesirable event once it occurs.

Secondary protection must be provided as a backup to the primary protection. Primary and secondary refer to levels of protection. While these levels are frequently provided by individual devices (e.g., pressure safety high (PSH)/pressure safety low, LSH/level safety low, PSV, and so forth), levels of protection can also be provided by other means. For example, the secondary level of protection required for a leak in a pressure vessel is the ESS (and not individual devices).

Undesirable events do not always cause injury, pollution, or damage, but they always have the potential to do so. To design a protection system to prevent or minimize injury, pollution, or damage, prevention efforts must be based on the assumption that undesirable events will cause these things.

11.3.14 Safety Analysis

Every process component can be grouped under one of the 10 process components listed in API *RP 14C*, [1] and process-related causes of injury, pollution, and damage can be grouped under one of the eight undesirable events. A safety analysis ties these two things together and is a tool for ensuring that a facility is protected fully. A safety analysis examines every process component in the facility to determine which undesirable events could be associated with each component, which safety devices are required for the protection of the component, and what responses the safety devices must make to ensure adequate protection. The three main components of a safety analysis are safety-analysis tables (SATs), safety-analysis checklists (SACs), and safety-analysis function evaluation (SAFE) charts.

11.3.14.1 Safety-Analysis Tables

SATs examine each process component as if it was standing alone. SATs consider each undesirable event that could possibly affect the component and then, for each undesirable event, lists associated causes, detectable abnormal conditions, and required locations for installing the protection devices. By examining each component as if it was standing alone, an adequate degree of protection can be determined for each particular component. When this is done for every component on the facility, the entire facility will be adequately protected. Verifying that each and every component is protected without considering other components ensures the greatest degree of consistent protection.

11.3.14.2 Safety-Analysis Checklists

There are times when the safety devices called for in SATs are not needed because engineering controls eliminate the need for a particular device. For example, the SAT calls for a PSV to protect a wellhead flowline from overpressure. However, if the MAWP of the flowline and associated equipment is greater than the maximum shut-in tubing pressure of the well, the component is already protected and the device is not needed.

A SAT-required safety device also no longer may be required if the same degree of protection is provided by another device located elsewhere. For example, if a PSV has been installed on an upstream flowline segment and if that upstream PSV provides an adequate degree of protection for the downstream flowline segment and its equipment, then a second PSV located on the downstream flowline segment is redundant.

SACs provide a guideline for eliminating redundant devices while maintaining the required level of protection. If it was not possible to eliminate redundant devices, production facilities would contain many more devices without gaining any additional protection. The time and expense of purchasing, installing, and maintaining redundant devices would be significant and unnecessary. It is important to realize that when a device can be eliminated, the device is eliminated and not the required level of protection. Two levels of protection will always be required. The SAC ensures that both levels of protection are maintained, with as few individual devices as possible. Figure 11.4 [1] shows an example of an SAT and an SAC for a flowline segment.

Undesirable Event	Cause	Detectable Abnormal Condition at Component
Overpressure	Blocked or restricted line Downstream choke plugged Hydrate plug Upstream flow-control failure Changing well conditions Closed outlet valve	High pressure
Leak	Deterioration Erosion Corrosion Impact damage Vibration	Low pressure

a. High-Pressure Sensor (PSH).

1. PSH installed.

2. Flowline segment has a MAWP greater than maximum shut-in pressure and is protected by a PSH on a downstream flowline segment.

b. Low-Pressure Sensor (PSL).

1. PSL installed.

2. Flowline segment is between the well and the first choking device and is less than 10 ft (3 m) in length or, in the case of an underwater installation, reasonably close to that distance.

c. Pressure Safety Valve (PSV).

1. PSV installed.

2. Flowline segment has a MAWP greater than the maximum shut-in pressure.

3. Two SDVs (one of which may be the SSV) with independent PSHs, relays, and sensing points are installed where there is adequate flowline volume upstream of any block valves to allow sufficient time for the SDVs to close before exceeding the MAWP.

4. Flowline segment is protected by a PSV on upstream segment.

5. Flowline segment is protected by a PSV on downstream component that cannot be isolated from the flowline segment, and there are no chokes or other restrictions between the flowline segment and the PSV.

d. Flow Safety Valve (FSV).

1. FSV installed.

2. Flowline segment is protected by FSV in final flowline segment.

Figure 11.4 Safety-analysis table and safety-analysis checklist for a flowline segment. *Reproduced courtesy of the American Petroleum Institute [1].*

11.3.14.3 Safety-Analysis Function-Evaluation Charts

SATs indicate which devices are needed on each component, and SACs determine which devices may be eliminated and what conditions must be met when eliminating the device. Neither SATs nor SACs indicate what the devices do or how the devices on one component relate to the devices on another component. SAFE charts are used to evaluate the function of each safety device and to document precisely what each safety device does. For example, the SAFE chart not only shows that a flowline PSH shuts off inflow, it indicates how it shuts off inflow (e.g., through the closing of a particular well's surface safety valve).

SAFE charts also indicate everything else that happens when a PSH trips. SAFE charts provide a mechanism for considering every component in the facility and then, for each component, fully account for each required safety device. SAFE charts are used to ensure that the facility is as fully protected as it should be and also can be used as a troubleshooting tool. For example, if a particular SDV keeps closing and nothing is out of range when investigated, the SAFE chart could be consulted to determine which specific devices cause the SDV to close. Each device then could be checked to determine which one is responsible for the SDV closures.

11.3.15 Manual Emergency Shutdown

The safety system should include features to minimize damage by stopping the release of flammable substances, de-energizing ignition sources, and shutting down appropriate equipment processes. This is accomplished by locating ESD stations at strategic locations to enable personnel to shut down the production facility. These ESD stations should be well marked and located conveniently (50–100 ft) from protected equipment, with back-up stations located some greater distance (250–500 ft) away. A good choice for location is along all exit routes. At least two widely separated locations should be selected.

The ESD can either shut down the entire facility, or it can be designed for two levels of shutdown. The first level shuts down equipment such as compressors, lean oil pumps, and direct fired heaters, and either shuts in the process or diverts flow around the process by closing inlet/outlet block valves and opening bypass valves. The second level shuts down the remaining utilities and support facilities, including generators and electrical feeds.

11.3.16 Annunciation Systems

These systems give early warning of impending trouble to allow personnel to take corrective action prior to a shut-in, and provide information about the initial cause of a shut-in. They are a vital part of any large shutdown system design. On smaller systems, process alarms may be minimal as there may not be sufficient time for personnel to react to the alarm before an automatic shutdown is initiated.

Annunciator panels should be in a central location with alarm annunciators and shutdown annunciators grouped separately. The first alarm and the first shut-down normally sound a horn and are annunciated. This is called “first-out indication.” Subsequent shutdown or alarm signals received by the panel are either not annunciated or are annunciated in a different manner so that the operator can determine the initiating cause of the process upset.

Alarm signals may come from the output signal used to control an operational valve. Shutdown signals should come from a completely separate instrument not dependent upon a normally used output signal for operation.

11.3.17 Conducting a Safety Analysis

The following steps comprise the process for conducting a production-facility safety analysis.

- Obtain an accurate process flow schematic (i.e., one that shows every process component as well as relevant operating parameters). Once the flow schematic is located, verify its accuracy because changes may have been made to the facility over a period of years that were not noted on the schematic. Verification involves walking around the facility to make sure that every process component located in the facility is pictured on the schematic. It also involves making sure that the flow schematic does not depict components that are no longer a part of the process facility and that maximum operating or working pressures are accurate. Failure to take this step jeopardizes the accuracy of both the flow schematic and the SAFE chart.
- Refer to each process component and the SATs to determine all required safety devices for each process component within the facility. Begin by referring to API *RP 14C*, Appendix A-1 through A-10. [1] Consult the SAT for each process component shown on the corrected flow schematic. Make sure each safety device called for in each component's section is shown on the schematic. Follow the example found in API *RP 14C*, Appendix E; that is, use “balloons” and ISA names for each device.

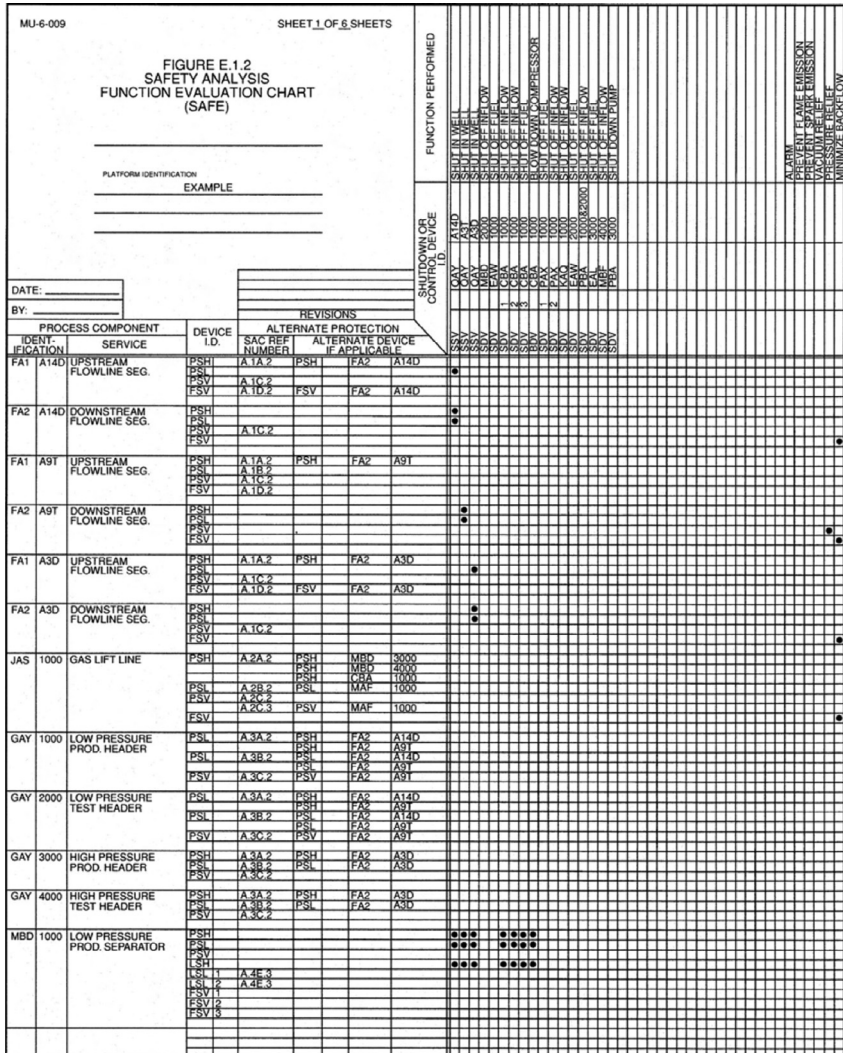
Before consulting the SAT for a particular component, it is important to first read everything written about that component in *API RP 14C*.

- Once each process component has been protected with the devices required by *API RP 14C*, [1] consult the SACs in *API RP 14C* to determine which, if any, devices provide redundant protection for each component. For each redundant device, make that device's solid-line balloon, which represents an installed safety device, into a dotted-line balloon, which represents an eliminated safety device. Remember, there will be adequate protection if there is an SAC reference number that applies to the situation. Look carefully at the descriptions following each SAC reference number, and determine if all the required conditions are met. If all the conditions are met, that particular device may be eliminated or the device may be left on the component. Remember, SACs allow for the elimination of redundant devices but do not require that they be eliminated. For those devices that will be eliminated, revise the schematic by replacing the solid-line balloon with a dotted-line balloon. See *API RP 14C*, Appendix E [1].
- Complete a SAFE chart for the facility; that is, fill out a blank SAFE chart with every component, safety device, and responding end device within the facility. Mark the SAFE chart to indicate the action taken by each safety device. In reality, completing a fresh, blank SAFE chart will seldom be required unless the initial safety analysis on a facility is being developed. Most often, an existing chart will be revised; however, knowing how to complete a fresh chart from scratch will make the job of revising an existing SAFE chart easier. Familiarity with SAFE charts enables them to be used to troubleshoot the facility.

The ability to complete a SAFE chart requires an understanding of how SAFE charts are arranged. SAFE charts are designed to be read horizontally and vertically. When read horizontally, the information pertains to all the process components within a facility plus their safety devices. When read vertically, the information pertains not only to the end devices affected by each safety device, but to their function as well (e.g., shut-in well, minimize backflow, and so forth). [Figure. 11.5](#) shows a typical SAFE chart [1].

11.3.18 Hazards Analysis

A hazards analysis identifies potential hazards, defines conditions necessary for each hazard, and identifies the source for each hazard. A hazard tree identifies potential hazards and determines the conditions necessary for these hazards to exist. A hazards analysis starts at the hazard tree's lowest level



(a)

Figure 11.5 (a) Safety-analysis function-evaluation chart (reproduced courtesy of the American Petroleum Institute [1]).

Continued

and attempts to break the path leading back to the hazard by eliminating one of the conditions.

Many of the sources and conditions identified on the hazard tree require considerations that have nothing to do with the way the process is designed, such as escape paths, electrical systems, fire-fighting systems, and insulation

MU-6-011		SHEET 3 OF 8 SHEETS		FUNCTION PERFORMED	
FIGURE E.1.2 (CONT'D) SAFETY ANALYSIS FUNCTION EVALUATION CHART (SAFE)		SHUTDOWN OR CONTROL DEVICE		FUNCTION PERFORMED	
PLATFORM IDENTIFICATION EXAMPLE		REVISIONS		FUNCTION PERFORMED	
DATE: _____		ALTERNATE PROTECTION		FUNCTION PERFORMED	
BY: _____		IF APPLICABLE		FUNCTION PERFORMED	
IDENTIFICATION	PROCESS COMPONENT	DEVICE I.D.	SAC REF NUMBER	ALTERNATE DEVICE I.D.	FUNCTION PERFORMED
MAJ 2000	GLYCOL FILTER	PSH A 4A.3	PSH	MBD 15000	SHUT DOWN COMPRESSOR
		PSL A 4B.4	PSV		SHUT DOWN COMPRESSOR
		PSV A 4C.2	PSV		SHUT DOWN COMPRESSOR
		LSH A 4D.3			SHUT DOWN COMPRESSOR
		LSL A 4E.2			SHUT DOWN COMPRESSOR
		FSV A 4F.2			SHUT DOWN COMPRESSOR
MBF 6000	SUCTION SCRUBBER	PSH			SHUT DOWN COMPRESSOR
		PSL			SHUT DOWN COMPRESSOR
		PSV			SHUT DOWN COMPRESSOR
		LSH			SHUT DOWN COMPRESSOR
		LSL			SHUT DOWN COMPRESSOR
		FSV 1 A 4F.2			SHUT DOWN COMPRESSOR
		FSV 2			SHUT DOWN COMPRESSOR
MBF 7000	INTERSTAGE SCRUBBER	PSH			SHUT DOWN COMPRESSOR
		PSL			SHUT DOWN COMPRESSOR
		PSV			SHUT DOWN COMPRESSOR
		LSH			SHUT DOWN COMPRESSOR
		LSL			SHUT DOWN COMPRESSOR
		FSV 1 A 4F.2			SHUT DOWN COMPRESSOR
		FSV 2			SHUT DOWN COMPRESSOR
MBF 7500	INTERSTAGE SCRUBBER	PSH A 4A.3	4 CBA 1000		SHUT DOWN COMPRESSOR
		PSL A 4B.3	4 CBA 1000		SHUT DOWN COMPRESSOR
		PSV A 4C.3	PSV 4 CBA 1000		SHUT DOWN COMPRESSOR
		LSH			SHUT DOWN COMPRESSOR
		LSL A 4E.3			SHUT DOWN COMPRESSOR
		FSV 1 A 4F.2			SHUT DOWN COMPRESSOR
		FSV 2			SHUT DOWN COMPRESSOR
MBF 8000	HIGH PRESSURE SCRUBBER	PSH A 4A.3	PSH MBD 3000		SHUT DOWN COMPRESSOR
		PSL	MBD 4000		SHUT DOWN COMPRESSOR
		PSV A 4B.3	PSH 6 CBA 1000		SHUT DOWN COMPRESSOR
		PSV	MBD 3000		SHUT DOWN COMPRESSOR
		LSH A 4D.2	PSL 6 CBA 1000		SHUT DOWN COMPRESSOR
		LSL A 4E.3			SHUT DOWN COMPRESSOR
		FSV 1			SHUT DOWN COMPRESSOR
		FSV 2			SHUT DOWN COMPRESSOR
MAF 1000	CONTACTOR	PSH A 4C.3	PSH MBD 3000		SHUT DOWN COMPRESSOR
		PSL	MBD 4000		SHUT DOWN COMPRESSOR
		PSV	PSH 6 CBA 1000		SHUT DOWN COMPRESSOR
		LSH A 4D.2			SHUT DOWN COMPRESSOR
		LSL A 4E.3			SHUT DOWN COMPRESSOR
		FSV 1 A 4F.2			SHUT DOWN COMPRESSOR
		FSV 2			SHUT DOWN COMPRESSOR
ABJ 1000	ATMOSPHERIC TANK	Vent			SHUT DOWN COMPRESSOR
		PSV			SHUT DOWN COMPRESSOR
		LSH			SHUT DOWN COMPRESSOR
		LSL			SHUT DOWN COMPRESSOR
ABJ 2000	GLYCOL SURGE TANK	Vent			SHUT DOWN COMPRESSOR
		PSV A 5B.3			SHUT DOWN COMPRESSOR
		LSH A 5C.4			SHUT DOWN COMPRESSOR
		LSL A 5D.2			SHUT DOWN COMPRESSOR
ABH 1000	SUMP	Vent			SHUT DOWN COMPRESSOR
		PSV			SHUT DOWN COMPRESSOR
		LSH			SHUT DOWN COMPRESSOR
		LSL A 5D.4			SHUT DOWN COMPRESSOR

(c)

Figure 11.5—Cont'd (c) Safety-analysis function-evaluation chart (reproduced courtesy of the American Petroleum Institute [1]).

Continued

procedures, testing, and drills, is required if the overall probability of any one chain leading to a hazard is to be acceptable. The different hazards analysis techniques can be applied at various stages during the course of the project to assess and mitigate potential hazards during design, construction, and operation of the facility.



Figure 11.5—Cont'd (d) Safety-analysis function-evaluation chart (*reproduced courtesy of the American Petroleum Institute [1]*).

Continued

11.3.19 Types of Hazards Analysis

Hazards analysis techniques fall in two broad categories. Some techniques focus on hazards control by ensuring that the design is in compliance with a preexisting standard practice. These techniques result from prior hazards analysis, industry standards and recommended practices, results of incident

recommended for most production facilities. The actual procedure by which the checklist is considered and the manner in which the evaluation is documented to ensure compliance varies from case-to-case.

The most common predictive technique used to analyze facilities that contain new equipment, or processes, or where there is an unusually high risk to personnel or the environment, is the Hazard and Operability technique or “HAZOP.” A HAZOP study requires a team of five to ten multidiscipline personnel consisting of representatives from engineering, operations, health, safety, and environmental staff. The facility is broken down into “nodes” (usually a major piece of equipment and its associated piping, valves, and instrumentation), and an experienced team leader guides the team through an analysis of each node using a predetermined list of “guide words” and “process parameters.” For example, the guide word “LOW” and process parameter “PRESSURE” results in questions being asked as to potential causes for lower than design pressure at the node. If the condition is possible, effects are analyzed and, if necessary, methods of mitigation are added until the risk is deemed acceptable. Although this method is time consuming, it proves to be a thorough method of analysis and is effective for a new process that has never been analyzed before or for a known process that incorporates new equipment. However, a checklist should be used in conjunction with a HAZOP to ensure that compliance with standard practice is not inadvertently overlooked by the HAZOP team.

11.3.20 Problems Commonly Encountered

There are several problem areas that seem to appear often in the results of hazards analyses. The most common are:

11.3.20.1 Relief Valve Sizes

Relief valves are often undersized for the required relieving rate, due either to poor initial design or changes in the process conditions that occurred during design. The most common system problem is that the relief valve was adequately sized for blocked discharge but not for the flowrate that could occur as a result of a failure in the open position of an upstream control valve (i.e., gas blowby).

11.3.20.2 Open and Closed Drains

Another common problem area is having open and closed drain systems tied together. Liquid that drains from pressure vessels “flash” at atmospheric pressures giving off gas. If this liquid flows in the same piping as open drains, the

gas will seek the closest exit to atmosphere it can find, causing a potential fire hazard at any open drain in the system.

Many accidents have occurred where gas has migrated through the drain system to an unclassified area where welding, or other hot work, was being performed.

11.3.20.3 Piping Specification Breaks

Piping pressure ratings should be designed so that no matter which valve is closed, the piping is rated for any possible pressure it could be subjected to, or is protected by a relief valve. When a spec break is taken from a higher to a lower MAWP, there must be a relief valve on the lower pressure side to protect the piping from overpressure. The relief valve can be either on the piping or, more commonly, on a downstream vessel. Spec break problems most commonly occur where a block valve exists on a vessel inlet, or where a bypass is installed from a high-pressure system, around the pressure vessel that has a relief valve, to a lower pressure system.

11.3.20.4 Electrical Area Classification

Another common mistake often uncovered is electrical equipment that is not consistent with the design area classification.



11.4 SAFETY MANAGEMENT SYSTEMS

A hazards analysis by itself cannot guarantee that an adequate level for safety is provided for a facility unless the hazard analysis is included as part of a comprehensive safety management system. In the United States every facility handling highly hazardous chemicals, including some onshore production facilities and most gas plants, must have a Process Safety Management (PSM) Plan in place. Offshore operators have developed a voluntary safety management system presented in API RP 75, "Recommended Practices for Development of a Safety and Environmental Management Program for Outer Continental Shelf (OCS) Operations and Facilities" (SEMP), which describes the elements that should be included in a safety management plan.

The requirements of both PSM and SEMP are, from a practical standpoint, identical and thus, SEMP can easily be applied to onshore facilities as well as offshore facilities. The basic concepts of SEMP are as follows:

11.4.1 Safety and Environmental Information

Safety and environmental information is needed to provide a basis for implementation of further program components such as operating procedures and hazards analysis. Specific guidelines as to what information is needed are contained in API RP 14J.

11.4.2 Hazards Analysis

This subject is addressed in the previous section of this chapter. Specific guidelines for performing hazards analysis are contained in API RP 14J.

11.4.3 Management of Change

Management of change is a program that helps to minimize accidents caused by changes of equipment or process conditions due to construction, demolition, or modification. Procedures should be set up to identify the various hazards associated with change. All changes, although sometimes minor, can result in accident and/or injury if proper steps are not implemented to make operators aware of the differences. Changes in facilities as well as changes in personnel should be managed to maintain the safety of all personnel and the environment.

11.4.4 Operating Procedures

The management program should include written facility operating procedures. These procedures should provide ample instruction for sound operation and be consistent with the safety and environmental information. Procedures should be reviewed and updated periodically to reflect current process operating practices. Procedures provide the means for educating new employees about the process and educating all employees about new equipment and practices.

11.4.5 Safe Work Practices

A disproportional amount of accidents occur during construction and major maintenance activities. Safe work practices are written with this in mind and, as a minimum, should cover the following:

- Opening of equipment or piping
- Lockout and tag-out of electrical and mechanical energy sources
- Hot work and other work involving ignition sources
- Confined space entry
- Crane operations

11.4.6 Training

Training for new employees and contractors, and periodic training of existing employees, is necessary to educate personnel to be able to perform their work safely and to be aware of environmental considerations. Training should address the operating procedures, the safe work practices, and the emergency response and control measures.

11.4.7 Assurance of Quality and Mechanical Integrity

Assurance of Quality and Mechanical Integrity of Critical Equipment

Procedures for assurance of quality in the design, fabrication, installation, maintenance, testing, and inspection for critical equipment are required. Safety requires that critical safety devices operate as intended and that process system components be maintained to be able to contain design pressures.

11.4.8 Pre-startup Review

A pre-startup safety and environmental review should be performed on all modified or newly constructed facilities.

11.4.9 Emergency Response and Control

An Emergency Action Plan should be established, assigning an emergency control center and appropriate personnel for emergency response. Drills should be carried out to ensure all personnel are familiar with these plans.

11.4.10 Investigation of Incidents

An investigation is required if an incident involving serious safety or environmental consequences or the potential for these consequences occurs. The purpose of such investigation is to learn from mistakes made and to provide corrective action. Investigations should be performed by knowledgeable personnel and should produce recommendations for safer working conditions.

11.4.11 Audit of Safety and Environmental Management Program Elements

Periodically, the SEMP elements should be audited to evaluate the effectiveness of the program. Auditing should be conducted by qualified personnel through interviews and inspections. If audits consistently find no deficiencies in the program, then management should conclude that the audit is not in itself being done properly, as there are always improvements that can be made in a safety management system.



11.5 SAFETY CASE AND INDIVIDUAL RISK RATE

The overall system for safety described above can be called the “API System.” It is based on a series of API Standards and Recommended Practices that can be summarized as a four-step system with each succeeding step encompassing the preceding steps:

1. Design and maintain a system for process upset detection and shut-down—RP 14C.
2. Design and select hardware with known reliability and mechanical integrity to contain pressure and mitigate failure consequences—all other API RP 14 series standards.
3. Follow system design concepts, documentation needs, and hazards analysis requirements—RP 14J.
4. Develop a management of safety system—RP 75.

This system has proven to provide adequate levels of safety in the Gulf of Mexico and other similar areas where it is possible to abandon the location during a catastrophic event. In the North Sea where harsh environmental conditions exist, a different approach to safety has evolved that is based on developing a safety case and calculating an individual risk rate (IRR) to show that the risk to any individual working in the facility is “as low as reasonably practicable,” or ALARP.

A safety case is a narrative that literally makes the case that an adequate level of safety has been reached for an installation. It requires looking at all potential hazards that could lead to a loss of the installation, a loss of life, or a major pollution event. A risk analysis is performed on each hazard evaluating the probability of the event occurring and describing the magnitude of the consequences. A discussion is then given of the measure undertaken to lower the probability of occurrence or to mitigate the consequences, and a “case” is made that the risk for the installation meets the ALARP safety criteria.

In the North Sea this is often done with detailed quantified risk assessments and the calculation of an overall IRR or risk of total loss of structure. Mitigation measures are incorporated until it can be shown that risk levels meet a minimum criteria *and* the cost of further mitigation has such high cost to benefit ratios that further mitigation is no longer “practicable.” These analyses tend to be rather long and complex and can negatively impact both project cycle time and cost. Indeed, as a check to ensure that basic known safety concepts are not inadvertently overlooked in the pile of documentation that is necessary for a safety case, the safety case approach should include

within it all the elements of the “API System.” Even if a safety case is performed, it is still necessary to ensure compliance with good practices and that all elements of a proper safety management system are included. Thus, a commonsense approach in the absence of government regulation is to use the API system for most installations and, in those instances where there is a large concentration of personnel or where abandoning the location may be impossible due to weather or remoteness, to use a qualitative safety case to think through fire-fighting and escape options.



11.6 RELIEF VALVES AND RELIEF SYSTEMS

11.6.1 Introduction

A relief system is an emergency system for discharging gas during abnormal conditions, by manual or controlled means or by an automatic pressure-relief valve from a pressurized vessel or piping system, to the atmosphere to relieve pressures in excess of MAWP. The relief system may include the relief device, the collection piping, flashback protection, and a gas outlet. A scrubbing vessel should be provided for liquid separation if liquid hydrocarbons are anticipated. The relief-system outlet may be either vented or flared. If designed properly, vent or flare emergency-relief systems from pressure vessels may be combined.

Some facilities include systems for depressuring pressure vessels in the event of an emergency shutdown. The depressuring-system control valves may be arranged to discharge into the vent, flare, or relief systems. The possibility of freezing and hydrate formation during high-pressure releases to the atmosphere should be considered.

There are three main engineering considerations when designing or modifying a relief system:

- Determining the relief requirements of individual pieces of equipment and selecting the appropriate devices to handle the imposed loads.
- Designing a relief header system that will handle the imposed loads or expansion modifications.
- Defining reasonable total relief loads for the combined relief header or disposal system and designing an appropriate disposal system with minimum adverse impact to personnel safety, plant-process system integrity, and the environment.

These considerations are interrelated in such a way that makes it impossible to establish a procedural guideline that would be valid for most cases. The design of one portion of a relief system must be considered in light of its effects on the relief system.

11.6.2 Relief Device Selection

11.6.2.1 Determining Individual Relief Loads

There are a number of industry codes, standards, and recommended practices that provide guidance in the sizing, selection, and installation of relief devices and systems. The American Society of Mechanical Engineers (ASME) Pressure Vessel Code, Sec. VIII, Division 1, Paragraph UG-127, lists the relief-valve code requirements [3]. API RP 520, Part 1, provides an overview of the types of relief devices, causes of overpressure, relief-load determination, and procedures for selecting and sizing relief devices [4]. API RP 520, Part 2, provides guidance on the installation of relief devices [5], and API RP 521 provides guidance on the selection and design of disposal systems [6].

11.6.2.2 Causes of Overpressure

The most common causes of overpressure in upstream operations are blocked discharge, gas blowby, and fire. When the worst-case relief load is caused by a control valve failing to open (blocked discharge), the relief device should be sized with full-sized trim in the control valve, even if the actual valve has reduced trim. When the worst-case relief load is caused by gas blowby, the relief device should be sized with full-sized trim in the smallest valve in the liquid-outlet line, even if the actual valve has reduced trim. Many vessels are insulated for energy savings. Thermal insulation limits the heat absorption from fire exposure as long as it is intact. It is essential that effective weather protection be provided so that insulation will not be removed by high-velocity fire-hose streams.

11.6.2.3 Types of Pressure-Relief Devices

The two primary types of relief devices are the relief valve and the rupture disk.

11.6.2.3.1 Relief Valves

The three basic types of pressure-relief valves are conventional spring loaded, balanced spring loaded, and pilot operated.

- Conventional spring loaded. In the conventional spring-loaded valve (Figure 11.6), the bonnet, spring, and guide are exposed to the released fluids. If the bonnet is vented to the atmosphere, relief-system backpressure decreases the set pressure. If the bonnet is vented internally to the outlet, relief-system backpressure increases the set pressure. The conventional spring-loaded valve is used in noncorrosive services and where backpressure is <10% of the set point.
- Balanced spring loaded. The balanced spring-loaded valve (Figure 11.7) incorporates a means to protect the bonnet, spring, and guide from the

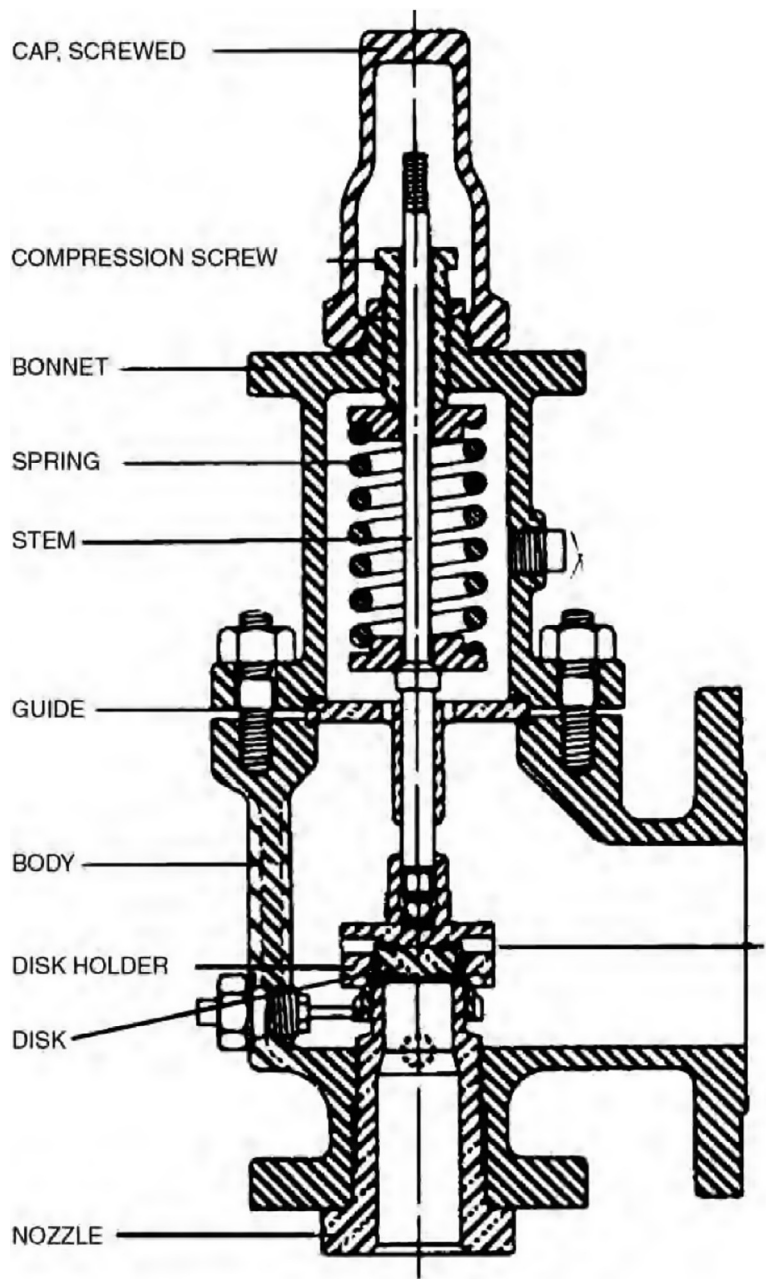


Figure 11.6 Conventional spring-loaded pressure-relief value.

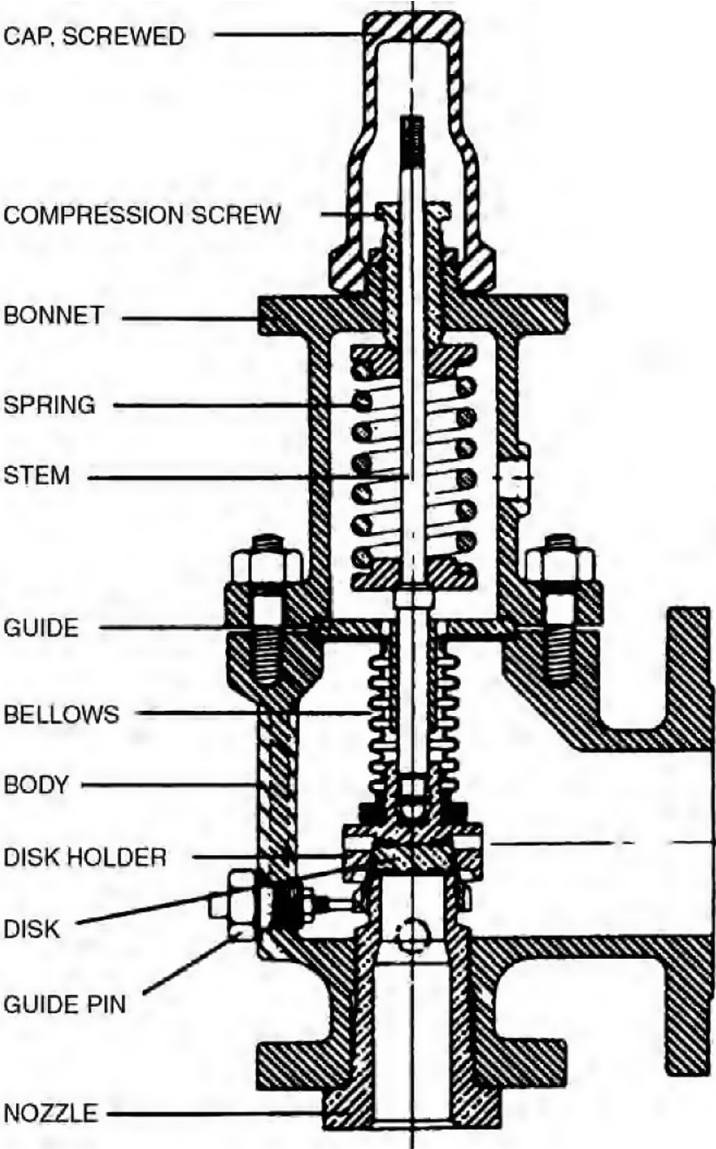


Figure 11.7 Balanced spring-loaded pressure-relief value. *Reproduced courtesy of the American Petroleum Institute [4].*

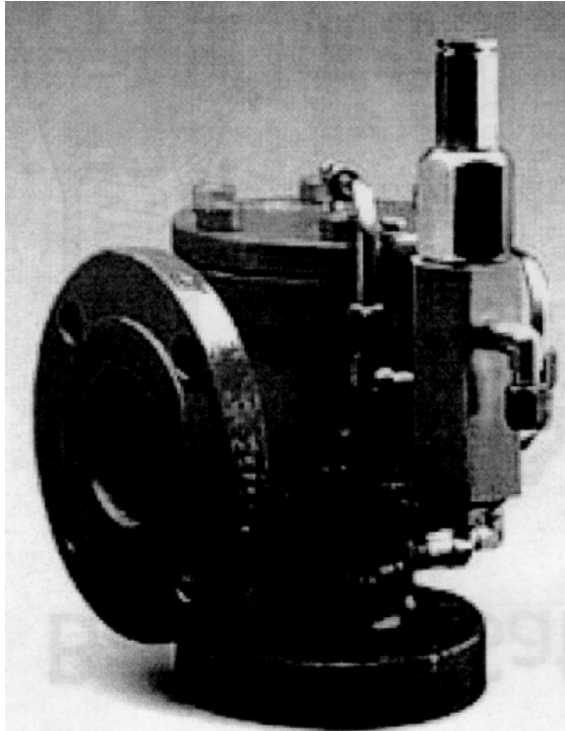


Figure 11.8 Pilot-operated pressure-relief valve. *Reproduced courtesy of the American Petroleum Institute [4].*

released fluids and minimizes the effects of backpressure. The disk area vented to the atmosphere is exactly equal to the disk area exposed to backpressure. These valves can be used in corrosive or dirty service and with variable backpressure.

- Pilot operated. The pilot-operated valve ([Figure 11.8](#)) is combined with and controlled by an auxiliary pressure pilot. The resistance force on the piston in the main valve is assisted by the process pressure through an orifice. The net seating force on the piston actually increases as the process pressure nears the set point.

11.6.2.3.2 Rupture-Disk Devices

The rupture-disk device ([Figure 11.9](#)) is a nonreclosing differential-pressure device actuated by inlet static pressure. The rupture disk is designed to burst at set inlet pressure. The device includes a rupture disk and a disk holder. The rupture disk may be used alone, in parallel with, or

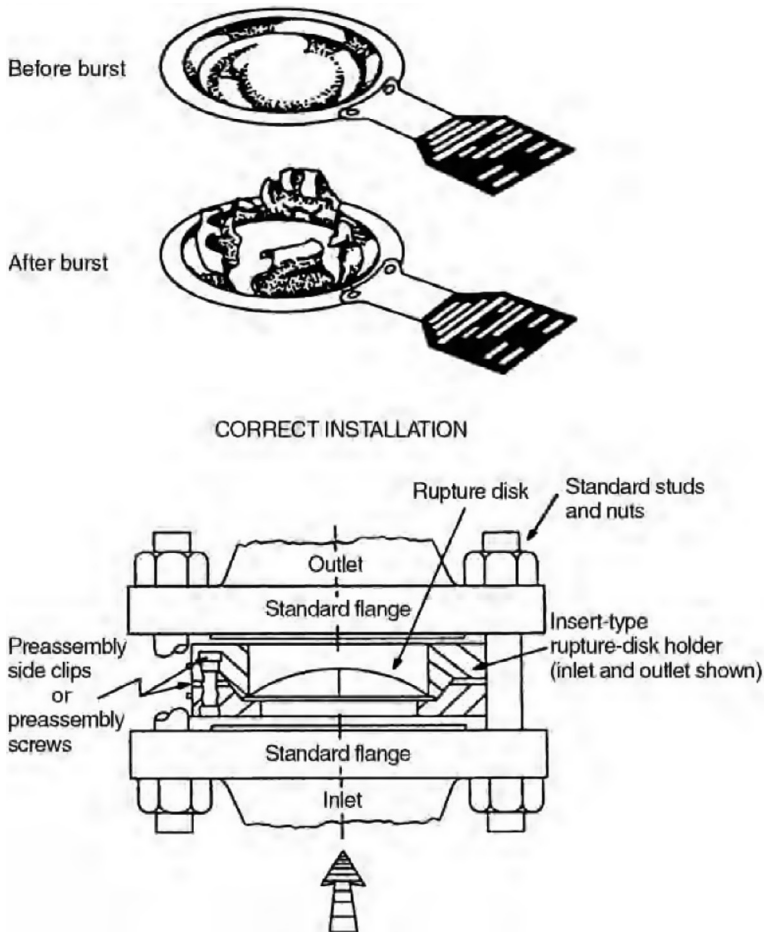


Figure 11.9 Conventional rupture disk. *Reproduced courtesy of the American Petroleum Institute [4].*

in conjunction with pressure-relief valves. They are manufactured in a variety of materials with various coatings for corrosion resistance.

11.6.3 Relief-System Considerations

The entire relief system must be considered before selecting the appropriate relief device. The relief headers should be designed to minimize pressure drop, thus allowing for future expansion and additional relief loads.

- Conventional spring-loaded-relief-valve considerations. Conventional valves require the relief header backpressure (superimposed plus

built up) to be $<10\%$ of the set pressure of the lowest-set relief valve tied into the header.

- **Balanced spring-loaded-valve considerations.** Balanced spring-loaded valves allow the use of smaller relief headers because of the larger pressure drops allowed, under maximum relief-flow conditions, as a result of higher allowable backpressure (40%). Balanced valves and relief headers are designed as a system to operate at a higher backpressure. The balanced valve is more expensive than conventional valves; however, the total cost of the use of balanced valves plus the smaller header system may be lower. Capacity is reduced at the larger backpressure, so it may not be the solution for all backpressure problems. In the bellows model, the bellows is a flexible pressure vessel that has a maximum backpressure limit that is lower in larger valve sizes. Bellows are available in a limited number of materials and may deteriorate rapidly under certain exposure conditions. Bellows should be checked periodically for leakage. A leaking bellows does not provide backpressure compensation, and it allows the relief header to leak to the atmosphere. The balanced valve commonly is used to tie a new low-pressure-relief load into an existing heavily loaded relief header or to protect the relief-valve top works from corrosive gases in the relief header.
- **Pilot-operated-valve considerations.** Pilot-operated valves should be considered for all clean services within their temperature limitations. They are well suited for pressures below 15 psig and are available with the pilot-pressure sensing line connected to either the valve inlet or to a different point. Pilot-operated valves provide tight shutoff with very narrow margins between operating pressure and set pressure.

11.6.4 Special Considerations

When selecting the appropriate relief devices to handle the imposed loads, several issues must be considered.

11.6.4.1 Set Pressure

Relief devices are normally set to relieve at the MAWP. The greater the margin between the set pressure and the operating pressure, the less likelihood there is of leakage. Aside from the requirements to compensate for superimposed backpressure, there is no reason to set a relief device at less than the MAWP.

11.6.4.2 Backpressure

The backpressure at the outlet of every relief device should be such that the device can handle its design capacity with the calculated backpressure under the design relief conditions.

11.6.4.3 Dual Relief Valves

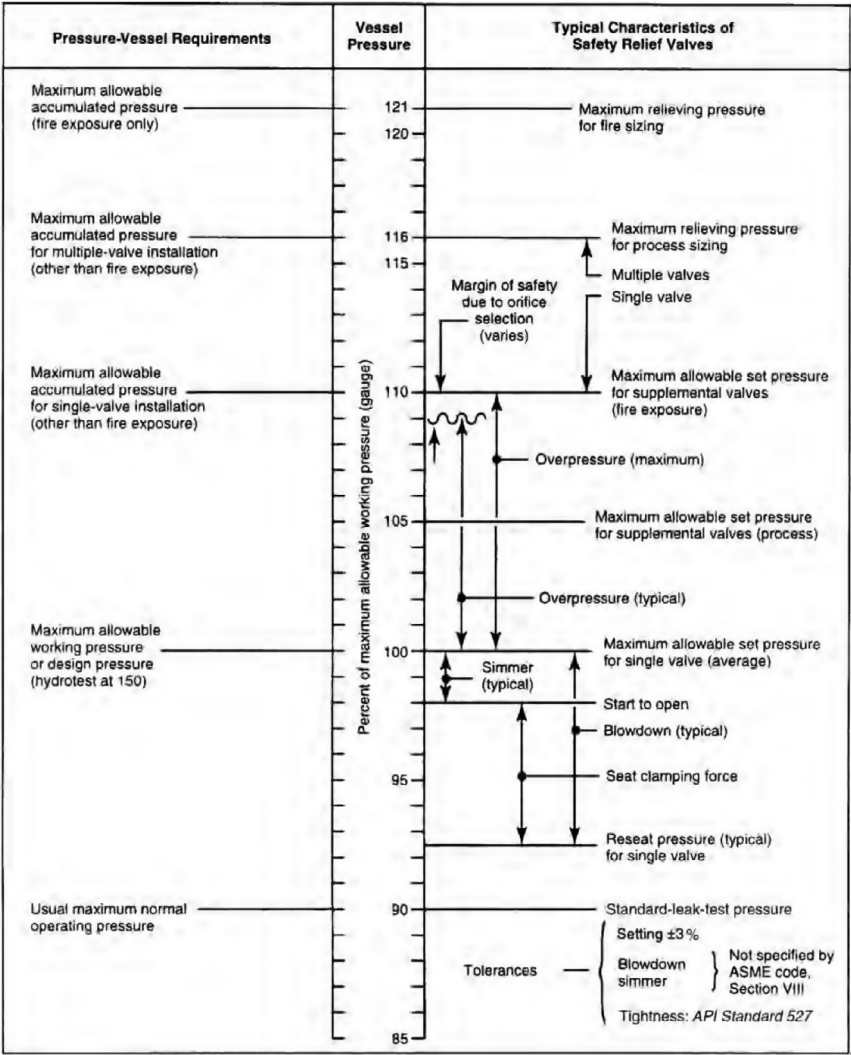
It is common practice to install two relief valves in critical process applications where a shutdown cannot be tolerated. The intent is that if the first relief valve lifts and fails to reseal, a second relief can be switched into service before the first valve is removed for maintenance, without shutting down or jeopardizing the process. This is accomplished by piping the relief valves in parallel and by putting a “car sealed” full-port ball or gate block valve on the inlet and outlet of each relief valve. One set of block valves is sealed open and the other sealed closed. ASME-approved selector valves are available that simplify relief-valve switching. This provides an interlock of parallel inlet and outlet block valves and ensures full protection for the process equipment.

11.6.4.4 Multiple Relief Valves

Multiple relief valves are required when the relief load exceeds the capacity of the largest available relief valve. It is good practice to install multiple relief valves for varying loads to minimize chattering on small discharges. ASME Sec. VIII, Division 1 [3], and API *RP 520*, Part 1 [4], (see [Figure 11.10](#)) both stipulate a 10% accumulation above the MAWP for a single relief valve and a 16% accumulation above the MAWP for multiple relief valves. The primary relief valve must be set at or below the MAWP. Supplemental relief valves should have staged pressures. The highest pressure may be set no higher than 105% above the MAWP. If different-sized relief valves are used, the smallest relief valve should be set to the lowest pressure.

11.6.4.5 Sizing the Relief Device

The most difficult factors for specifying a relief device are determining the limiting cause of pressure relief, determining the relief load and properties of the discharge fluid, and selecting the proper relief device. When the loads are known, the sizing steps are straightforward. API *RP 520*, Part 1, provides formulas for determining the relief-valve orifice area for vapor, liquid, and steam relief [4]. [Figure 11.11](#) shows standard orifices available by letter designation, orifice area, and body size. The size of a relief valve should be checked for the following conditions.



- Notes:
- 1. The operating pressure may be any lower pressure required.
 - 2. The set pressure and all other values related to it may be moved downward if the operating pressure permits.
 - 3. This figure conforms with the requirements of Section VIII, Division 1, of the ASME Code.
 - 4. The pressure conditions shown are for safety relief valves installed on a pressure vessel (vapor phase).

Figure 11.10 Pressure levels of pressure-relief values. Reproduced courtesy of the American Petroleum Institute [4].

		ORIFICE AREA, IN. ²										
STANDARD ORIFICE DESIGNATION	D	0.110	•	•	•							
	E	0.196	•	•	•							
	F	0.307	•	•	•							
	G	0.503			•	•	•					
	H	0.785				•	•					
	J	1.287					•	•	•			
	K	1.838						•				
	L	2.853						•	•			
	M	3.60							•			
	N	4.34							•			
	P	6.38							•			
	Q	11.05								•		
	R	16.0								•	•	
	T	26.0										•
		1 × 2	1.5 × 2	1.5 × 2.5	1.5 × 3	2 × 3	2.5 × 4	3 × 4	4 × 6	6 × 8	6 × 10	8 × 10
VALVE BODY SIZE (INLET DIAMETER × OUTLET DIAMETER), IN.												

Figure 11.11 Pressure-relief-value orifice designations.

11.6.4.5.1 Blocked Discharge

One design condition for the sizing of a relief valve is to assume that it must handle the total design flow rate (gas plus liquid) into the component. It is possible to isolate a process component or piping segment for maintenance by blocking all inlets and outlets. On startup, all outlet valves could be left closed inadvertently. If the inlet source can be at a higher pressure than the MAWP of the process component, only a properly sized relief valve could keep the process component from rupturing as a result of overpressure.

11.6.4.5.2 Gas Blowby

On tanks and low-pressure vessels normally receiving liquids from higher-pressure upstream vessels, the maximum flow rate through the relief valve often is determined by gas blowby. This situation occurs when the level controller or level control valve of the upstream vessel fails in the open position or a drain valve from an upstream vessel fails in the open position, allowing liquid and/or gas to flow into the component evaluated. Under blowby conditions, both the normal liquid and gas outlets on the component being evaluated are functioning properly. However, the gas flow into the component could greatly exceed the capacity of the normal gas outlet. This excess gas flow must be handled by the relief valve to keep from exceeding the component’s MAWP. Gas-blowby conditions also can occur when a

pressure regulator feeding a component fails in the open position, creating a higher than designed inlet flow rate of gas.

Gas-blowby rate is the maximum that can flow given the pressure drop between the upstream component and the component being evaluated. In computing the maximum rate that can flow because of pressure drop, consideration should be given to the effects of control valves, chokes, and other restricted orifices in the line. A more conservative approach would be to assume that these devices have been removed or have the maximum-sized orifice that could be installed in the device.

11.6.4.5.3 Fire/Thermal Expansion

The pressure in process components exposed to the heat from a fire will rise as the fluid expands and the process liquid vaporizes. For tanks and large low-pressure vessels, the need to vent the liberated gas may govern the size of the vent or relief valve. Fire sizing a relief valve only keeps pressure buildup to <120% of the MAWP. If the component is subjected to a fire for a long time, it may fail at a pressure less than the MAWP because a metal's strength decreases as temperature increases.

On components that can be isolated from the process, it is possible for the process fluid contained in the component to be heated. This is especially true for cold (relative to ambient) service or when the component is heated (such as a fired vessel or heat exchanger). It is also true for compressor cylinders and cooling jackets. The relief valves on such components should be sized for thermal expansion of the trapped fluids. This normally will not govern the final size selected unless no relief valve is needed for the other conditions.

11.6.5 Installation Considerations

The installation of a relief device requires careful consideration of the inlet piping, pressure-sensing lines (where used), and startup procedures. Poor installation may render the relief device inoperable or severely restrict the valve's relieving capacity. Either condition compromises the safety of the facility. Many relief-valve installations have block valves before and after the relief valve for in-service testing or removal; however, these block valves must be car sealed or locked open.

11.6.5.1 Inlet Piping

API *RP 520*, Part 2 [5], and ASME code [3] limit the inlet pressure loss to a PSV of 3% of set pressure where the pressure loss is the total of the inlet loss, line loss, and block-valve loss (if used). Loss is calculated with the maximum

rated flow through the relief valve. To minimize the inlet pressure drop to a relief valve, a conservative guideline is to keep the equivalent length-to-diameter ratio of the inlet piping to the relief valve at 5 or less. For pressure-drop limitations and typical piping configurations, refer to API *RP 520*, Part 2 [5].

11.6.5.2 Discharge Piping

The discharge piping should be designed so that the backpressure does not exceed an acceptable value for any relief valve in the system. Piping diameters generally should be larger than the valve-outlet size to limit backpressure. Lift and set pressures of pilot-operated relief valves with the pilot vented to the atmosphere are not affected by backpressure; however, if the discharge pressure can exceed the inlet pressure (e.g., tanks storing low-vapor-pressure material), a backflow preventer (vacuum block) must be used. The set pressure for balanced spring-loaded relief valves will not be as affected by backpressure as conventional spring-loaded relief valves are. Balanced relief valves will suffer reduced lift as backpressure increases.

11.6.5.3 Reactive Forces

On high-pressure valves, the reactive forces during relief are substantial and external bracing may be required. Refer to the formulas in *RP 520*, Parts 1 [4] and 2 [5] for computing these forces.

11.6.5.4 Tailpipe Considerations

Relief valves that are not connected to a closed relief system should have tailpipes to direct the relieving gases to a safe area away from personnel. The tailpipe should be sized for a maximum exit velocity of 500 ft/s. This ensures that the gas/air mixture is below the lower flammable limit or lower explosive limit at ~120 pipe diameters away from the tailpipe. Tailpipes should be supported at the bottom of the elbow. A small hole or a “weep hole” (minimum of ¼ in. in diameter) should be installed in the bottom of the elbow to drain liquids that enter through the tailpipe opening. The weep hole should be pointed away from process components, especially those classified as an ignition source.

11.6.5.5 Rapid Cycling

Rapid cycling can occur when the pressure at the valve inlet decreases at the start of the relief valve flow because of excessive pressure loss in the piping upstream of the valve. Under these conditions, the valve will cycle rapidly, a

condition referred to as “chattering.” Chattering is caused by the following sequence. The valve responds to the pressure at its inlet. If the pressure decreases during flow below the valve reseal point, the valve will close; however, as soon as the flow stops, the inlet-pipe pressure loss becomes zero and the pressure at the valve inlet rises to vessel pressure once again. If the vessel pressure is still equal to or greater than the relief-valve set pressure, the valve will open and close again. An oversized relief valve may also chatter because the valve may quickly relieve enough contained fluid to allow the vessel pressure to momentarily fall back to below set pressure, only to rapidly increase again. Rapid cycling reduces capacity and is destructive to the valve seat in addition to subjecting all the moving parts in the valve to excessive wear. Excessive backpressure also can cause rapid cycling, as discussed previously.

11.6.5.6 Resonant Chatter

Resonant chatter occurs when the inlet piping produces excessive loss at the valve inlet and the natural acoustical frequency of the inlet piping approaches the natural frequency of the valve’s moving parts. The higher the set pressure, the larger the valve size, or the greater the inlet-pipe pressure loss, the more likely resonant chatter will occur. Resonant chatter is uncontrollable, that is, once started it cannot be stopped unless the pressure is removed from the valve inlet. In actual practice, the valve can break down before a shutdown can take place because of the very large magnitude of the impact force involved. To avoid chattering, the pressure drop from the vessel nozzle to the relief valve should not exceed 3% of the set pressure. *RP 520*, Part 2 covers the design of relief-valve inlet piping [5]. Pilot-operated relief valves with remote sensing pilots can operate with higher inlet-piping pressure drops.

11.6.5.7 Isolation (Block) Valves

There is no industry standard or RP for isolation valves, and practices vary widely. Installed isolation block valves allow the testing of spring-loaded relief valves in place, thus eliminating the need to remove the vessel from service while bench testing the relief valve, and allow the relief device to be isolated from the closed relief system when performing maintenance and repair. The ASME Unfired Pressure Vessel Code allows the use of isolation valves below relief valves [3]. *ASME Pressure Vessel Code*, Appendix M, describes special mandatory requirements for isolation valves. The ASME Boiler Code [3] prohibits them, and the Occupational Safety and

Health Administration [7] prohibits them on instrument air receivers. Because improper use of an isolation valve may render a relief valve inoperative, the design, installation, and management of these block valves should be evaluated carefully to ensure that plant safety is not compromised. See *RP 520*, Part 2, for typical block-valve installations under relief valves [5].

11.6.5.8 Relief-Valve Configurations

There is no industry standard or RP that addresses this topic. Some of the more common relief-valve configurations are listed here and are shown in Figure 11.12.

- Installation of full open isolation (block) valves upstream and downstream of relief valves. Isolation valves should be car sealed open (locked open), and a log should be kept. These valves should be discouraged where the potential overpressure is twice the maximum allowable pressure. A test connection should be provided on all spring-loaded relief valves. The installation of two relief valves (100% redundant) should be considered so that one relief valve can be left in service at all times.

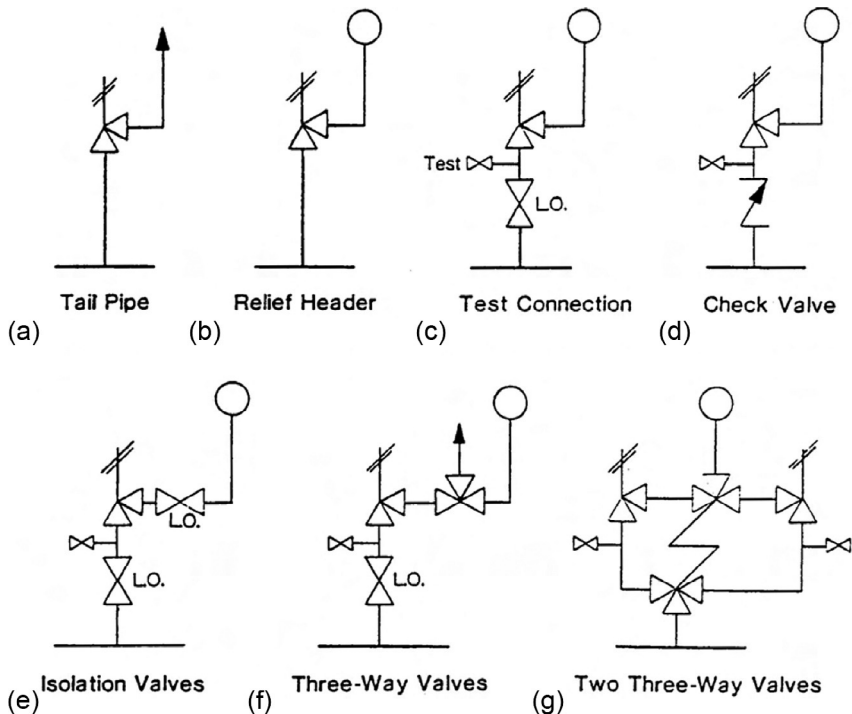


Figure 11.12 Relief-valve configurations.

- Installation of pilot-operated valves without isolation valves. This configuration allows for the testing of pilot set pressure only and requires full plant shut-in for relief-valve repair and maintenance.
- Installation of three-way valves with one port open to a tailpipe or a vent stack. This configuration allows for valve maintenance and repair without requiring plant shut-in and ensures a path to the atmosphere if the three-way valve is left in the wrong position.
- Installation of two two-way valves, connected by mechanical linkage, and two relief valves. This configuration provides all the advantages of isolation valves. In addition, it is impossible to isolate a process component by mistake. The only disadvantage of this configuration is the initial cost.
- Installation of a check valve in lieu of an isolation valve. This configuration is not allowed by the *ASME Pressure Vessel Code* because the check valve may fail or cause excessive pressure drop [3].

11.6.5.9 Guidelines for Determining the Number of Relief Devices

There is no industry standard or RP for determining the number of relief devices, and installations vary widely. Sometimes there are two relief devices (100% standby) on vessels receiving production directly from the wells. The primary relief valve is set at MAWP. If the second relief device is another relief valve, the set pressure of the second relief valve is set 10% above the primary relief valve. If the second relief device is a rupture disk (entirely redundant against all possible relieving scenarios), the pressure is set at 15–25% above the primary relief device. This setting ensures that the rupture disk will not rupture when the design primary relieving rate is reached at the set pressure plus 10% overpressure. Primary and standby relief rates are considered adequate for fire sizing.

Some companies install two relief valves on all critical installations so that plant shutdowns are not required during testing and maintenance. If the secondary relief device is being relied upon to provide any portion of any required relieving capacity (blocked discharge, gas blowby, fire, and so forth), then the secondary device should be set in accordance with the rules of *RP 520*, Parts 1 [4] and 2 [5], (i.e., ASME Sec. VIII, Division 1, paragraph UG-134) [3].

11.6.5.10 Liquid-Discharge Considerations

Condensed mists have liquid droplets that are $<20\text{--}30\text{ }\mu\text{m}$ in diameter. Testing and experience have shown that with a slight wind, the envelope of

flammability for this type of mist is the same as that for a vapor. Liquids will settle to grade, thus presenting a fire and pollution hazard; therefore, the relief device should be installed in the vapor space of process vessels with an LSH that alarms and shuts in flow when activated. The LSH should be set no higher than 15% above the maximum operating level, while the relief valve should be set no higher than the MAWP of the process component. Scrubbers and knockout drums should be installed in flare, vent, and relief lines to separate and remove liquid droplets from the discharge.



11.7 FLARE AND VENT DISPOSAL SYSTEMS

11.7.1 Disposal-System Design

A flare or vent disposal system collects and discharges gas from atmospheric or pressurized process components to the atmosphere to safe locations for final release during normal operations and abnormal conditions (emergency relief). In vent systems, the gas exiting the system is dispersed in the atmosphere. Flare systems generally have a pilot or ignition device that ignites the gas exiting the system because the discharge may be either continuous or intermittent. Gas-disposal systems for tanks operating near atmospheric pressure are often called atmospheric vents or flares, and gas-disposal systems for pressure vessels are called pressure vents or flares. A flare or vent system from a pressurized source may include a control valve, collection piping, flashback protection, and a gas outlet. A scrubbing vessel should be provided to remove liquid hydrocarbons. A flare or vent system from an atmospheric source may include a pressure-vacuum valve, collection piping, flashback protection, and a gas outlet. The actual configuration of the flare or vent system depends on the hazards assessment for the specific installation.

API RP 520, Part 1, Sec. 8 [4], and *RP 521*, Secs. 4 and 5 [6], cover disposal and depressuring system design. *API RP 521*, Appendix C, provides sample calculations for sizing a flare stack. *API RP 521*, Appendix D, shows a flare-stack seal drum, a quench drum, and a typical flare installation [6].

11.7.2 Knockout Drums

API RP 521, Paragraph 5.4.2, provides detailed guidance for the design of knockout drums (also called relief drums or flare or vent scrubbers) [6]. All flare, vent, and relief systems must include a liquid knockout drum. The knockout drum removes any liquid droplets that carry over with the gas relief sent to the flare. Most flares require that the particle size be reduced

to a minimum of $<300\text{ }\mu\text{m}$. API *RP 14J* suggests sizing for liquid droplets between 400 and 500 μm [2]. Most knockout drums are horizontal with a slenderness ratio (length-to-diameter ratio) between 2 and 4. A horizontal knockout drum must have a diameter large enough to keep the vapor velocity low enough to allow entrained liquids to settle or drop out.

Knockout drums operated at atmospheric pressure should be sized to handle the greatest liquid volume expected at the maximum rates of liquid buildup and pump out. API *RP 521* suggests 20–30 min of liquid holdup [6]. This is not practical in upstream operations. In onshore operations, it is recommended to take 20% of the maximum potential liquid stream and provide a 10-min liquid holdup. For offshore operations, it is recommended to provide normal separation-retention times (1–3 min on the basis of API gravity) and an emergency dump design to handle the maximum liquid flow with no valves. An emergency sump (disposal) pile is recommended to dispose of the liquid, and a seal in the pile is recommended to contain the backpressure in the drum.

Knockout drums normally are operated at atmospheric pressure. To maintain an explosion, the MAWP of the knockout drum usually is set at 50 psig. Stoichiometric hydrocarbon/air explosions produce peak pressures seven to eight times the normal pressure.

11.7.3 Flashback Protection

Flashback protection (the possibility that the flame will travel upstream into the system) should be considered for all disposal systems because flashback can result in pressure buildup in upstream piping and vessels. Flashback is more critical where there are tanks or pressure vessels with a MAWP <125 psig and in flare systems. *RP 520* discusses flashback protection for pressure vents and flares [4], and API *RP 2000* discusses atmospheric vents and flares [8]. API *RP 14C* recommends that vents from atmospheric vessels contain a flame arrestor [1]. Because the flame arrestor can plug, a secondary pressure/vacuum valve without a flame arrestor should be considered for redundancy. The secondary system should be set at a pressure high enough and vacuum low enough so that it will not operate unless the flame arrestor on the primary system is plugged.

Pressure vents with vessels rated 125 psig and above normally do not need flashback protection. In natural-gas streams, the possibility of vent ignition followed by flash backpressures above 125 psig is minimal. When low-pressure vessels are connected to pressure vents, molecular or fluidic seals

and purge gas often are used to prevent flashback. If relief valves are tied into the vent, the surge of flow when a relief valve opens could destroy a flame arrestor and lead to a hazardous condition. Also, there is a potential for flame arresters to become plugged. A means of flame snuffing should be considered for vent systems.

Flares have the added consideration of a flame always being present, even when there is a very low flow rate. They are typically equipped with molecular or fluidic seals and a small amount of purge gas to protect against flashback.

11.7.4 Seal Drums

Knockout drums are sized with the gas-capacity equations referred to in Chapter 4 of Volume 1 of the series. Liquid seal drums are vessels that are used to separate the relief gases and the flare/header stack by a layer of liquid. Water (or water/glycol mixture) is normally the sealing fluid. The flare gas (or purge gas) is forced to bubble through a layer of water before it reaches the flare stack. This prevents air or gas from flowing backward beyond the water seal. Seal drums serve as a final knockout drum to separate liquid from the relief gases.

In a deep seal drum, the depth of the sealing fluid is designed to be equal to the staging pressure of the staged flare system. The sealing-fluid depth in most staging seal drums is typically in the range of 2-5 psig, which is equivalent to 5-12.5 ft of water column. In a shallow seal drum (conventional flashback prevention), the water seals have only a 6-10-in. water-column depth. It is important to design the deep seal drum with a proper gas velocity at the staging point to ensure that all the sealing fluid is displaced quickly at the staging pressure (an effect similar to a fast-acting valve actuator). It is also common to design the deep seal drum with a concentric overflow chamber to collect the displaced sealing fluid. The overflow chamber can be designed to flow back automatically into the sealing chamber once the gas velocity decreases below the rate required for closing off the second stage.

The depth of the liquid seal drum must be considered in calculating the relief-header backpressure. This depth is set by the flare supplier, but it usually can be altered somewhat, with the supplier's concurrence, to suit plant conditions. Typical seal depths are 2 ft for elevated flares and 6 in. for ground flares. The height of the liquid seal can be determined by

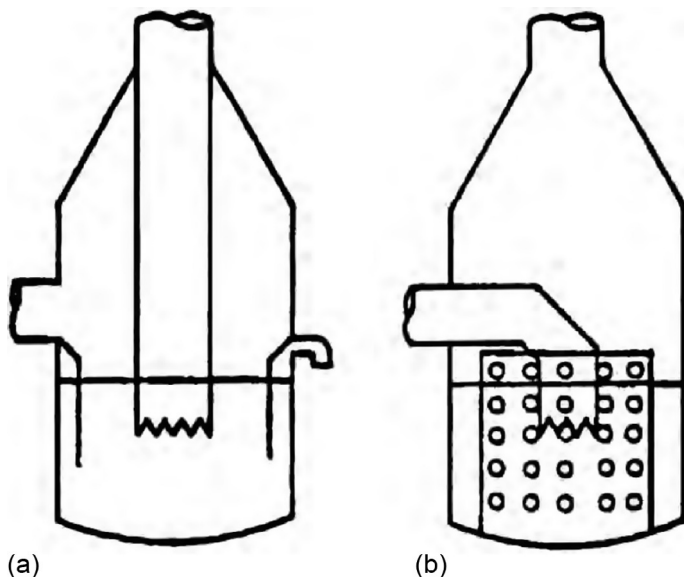


Figure 11.13 Seal-drum configuration with (a) Displacement Seal and (b) Perforated Antislosh baffle.

$$h = \frac{(144)p}{\rho} \quad (11.1)$$

where h = height of liquid seal, p = maximum allowable header back-pressure, and ρ = sealing-liquid density.

The vessel-free area for gas flow above the liquid level should be a minimum of 3 ft or three times the inlet pipe cross-sectional area to prevent surges of gas flow to the flare and to provide space for disengagement.

API RP 521 states that surging in seal drums can be minimized with the use of V-notches on the end of the dip leg [6]. If the water sloshes in the seal drum, it will cause pulsations in the gas flow to the flare, resulting in noise and light disturbances. Thus, most facilities prefer either a displacement seal or a perforated antislosh baffle. [Figure 11.13](#) shows seal-drum configurations.

11.7.5 Molecular Seals

Molecular seals cause flow reversal. They normally are located below the flare tip and serve to prevent air entry into the stack. Molecular seals depend on the density difference between air and hydrocarbon gas. Light gas is trapped at the top of the U-tube. A continuous stream of purge gas is required for proper functioning of the gas seal, but the amount of purge

gas is much less than would be required without the seal. The main advantages over liquid seals are that they do not slosh and they produce much less oily water. Gas seal must be drained, and the drain loop must be sealed. Because a gas seal with an elevated flare is required to keep air out of the flare stack, the liquid seal usually is omitted from an elevated-only flare system. If a vapor-recovery compressor is used, a liquid seal is used to provide a minimum header backpressure.

11.7.6 Fluidic Seals

Fluidic seals are an alternative to gas seals. Fluidic seals use an open wall-less venturi, which permits flow out of the flare in one direction with very little resistance but strongly resists counter-flow of air back into the stack. The venturi is a series of baffles, like open-ended cones in appearance, mounted with the flare tip. The main advantages of fluidic seals are that they are smaller, less expensive, and weigh less, and thus have less structural load on the flare stack than molecular seals. However, fluidic seals require more purge gas than molecular seals.

11.7.7 Flame Arrestors

Flame arrestors are used primarily on atmospheric vents and are not recommended on pressurized systems. Because of the acceleration of the flame, the flame arrestor must be installed ~ 10 pipe diameters from the exit, which prevents the flame from blowing through the arrestor. The length of the tube and surface area provided keep the metal cool. The major drawbacks of flame arrestors are that they are easily plugged, can become coated with liquid, and may not be strong enough for pressure-relief systems.

11.7.8 Flare Stacks

API RP 521, Sec. 5.4.3, covers the design of elevated flares [6]. *API RP 521*, Appendix C, provides examples of full designs of a flare stack [6]. Most flares are designed to operate on an elevated flare stack or on angled booms on offshore platforms.

11.7.9 Elevated-Flare-Stack Designs

Figure 11.14 shows an example of an elevated-flare-stack design.

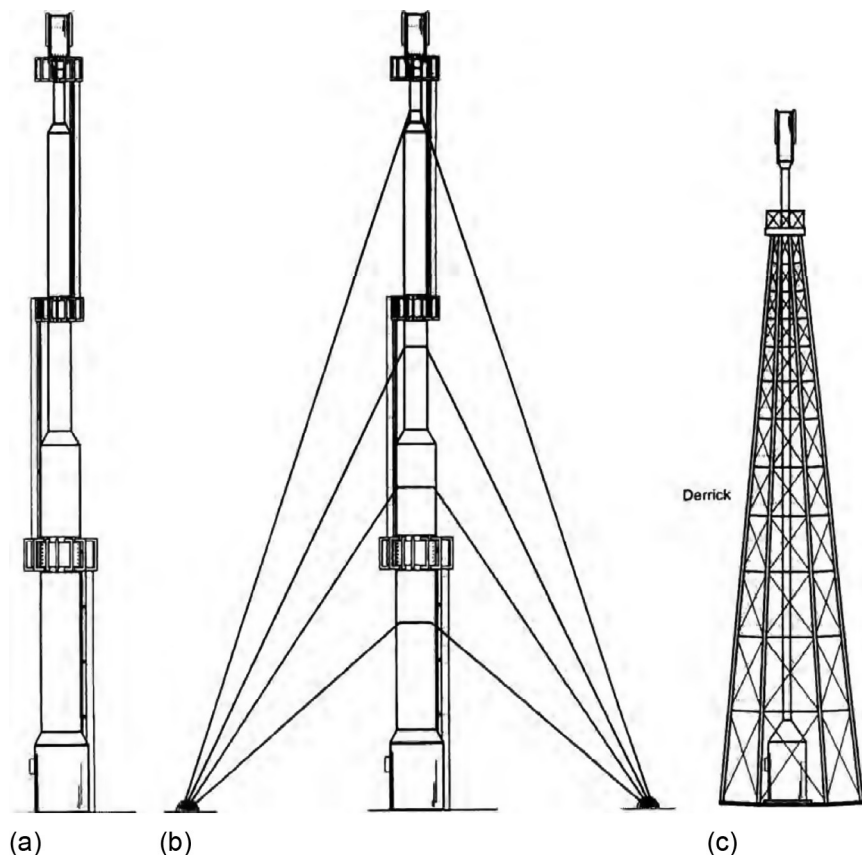


Figure 11.14 Elevated flare-stack configurations: (a) Self-Supported, (b) Guyed Supported, and (c) Derrick Supported.

11.7.9.1 Self-Supported Stacks

This is the simplest and most economical design for applications requiring short-stack heights (up to 100 ft overall height); however, as the flare height and/or wind loading increases, the diameter and wall thickness required become very large and expensive.

11.7.9.2 Guy-Wire-Supported Stacks

This is the most economical design in the 100–350-ft height range. The design can be a single-diameter riser or a cantilevered design. Normally, sets of three wires are anchored 120 degrees apart at various elevations (1–6).

11.7.9.3 Derrick-Supported Stacks

This is the most feasible design for stack heights above 350 ft. They use a single-diameter riser supported by a bolted framework of supports. Derrick supports can be fabricated from pipe (most common), angle iron, solid rods, or a combination of these materials. They sometimes are chosen over guy-wire-supported stacks when a limited footprint is desired.

11.7.10 Offshore Flare-Support Structures

Because offshore production platforms process very large quantities of high-pressure gas, the relief systems and, therefore, the flare systems, must be designed to handle extremely large quantities of gas quickly. By nature, flares normally have to be located very close to production equipment and platform personnel or located on remote platforms. Maximum emergency-flare design is based on emergency shut-in of the production manifold and quick depressurization of the system. Maximum continuous-flare design is based on loss of produced-gas transport, single compression shutdown, gas-turbine shutdown, and so forth. Typical flare mountings on an offshore platform are angled boom mounting (most common), vertical towers, or remote flare platforms. [Figure 11.15](#) shows typical offshore flare-support structures.

Selection of the flare structure depends on such factors as water depth, the distance between the flare and the production platform, relief gas quantity, toxicity, allowable loading on the flare structure, location of personnel, location of drilling derrick, locations of adjacent platforms, and whether the flaring is intermittent or continuous.

11.7.10.1 Flare Booms

Flare booms extend from the edge of the platform at an angle of 15–45° and are usually 100–200 ft long. Sometimes two booms oriented 180° from one another are used to take advantage of prevailing winds. [Figure 11.16](#) shows a diagram of an offshore flare boom.

11.7.10.2 Derrick-Supported Flares

Derrick-supported flares (see [Figure 11.17](#)) are the most common flare towers used offshore. They provide the minimum footprint (four-legged design) and dead load, which are critical design parameters for offshore flares and normally are used when space is limited and relief quantities are moderate. Disadvantages of derrick-supported flares include possible crude-oil spill onto the platform, interference with helicopter landing, and higher radiation intensities.

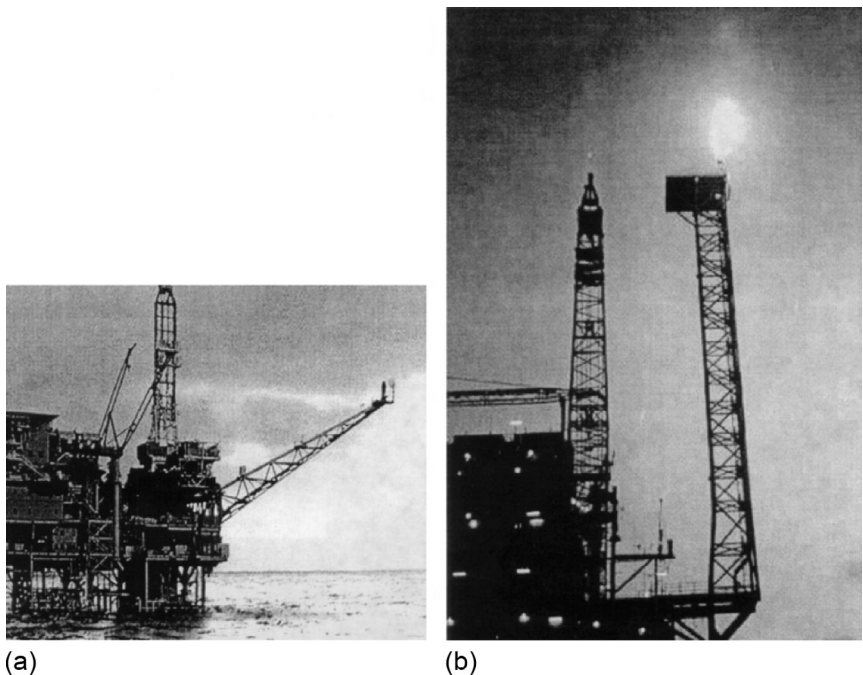


Figure 11.15 Typical offshore flare-support structures: (a) Angle Flare Boom and (b) Vertical Tower.

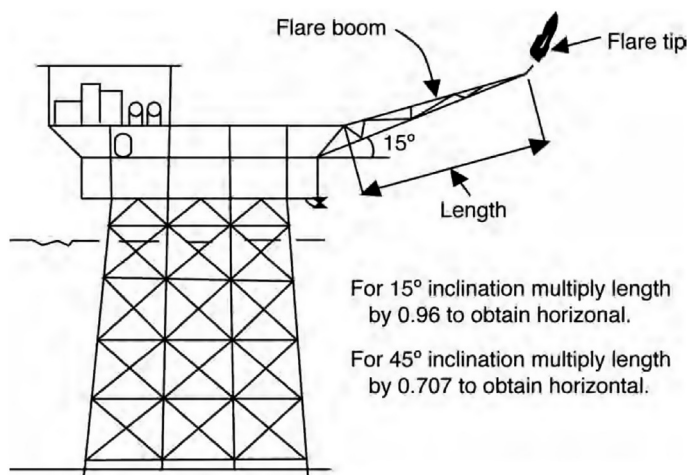


Figure 11.16 Offshore flare boom.

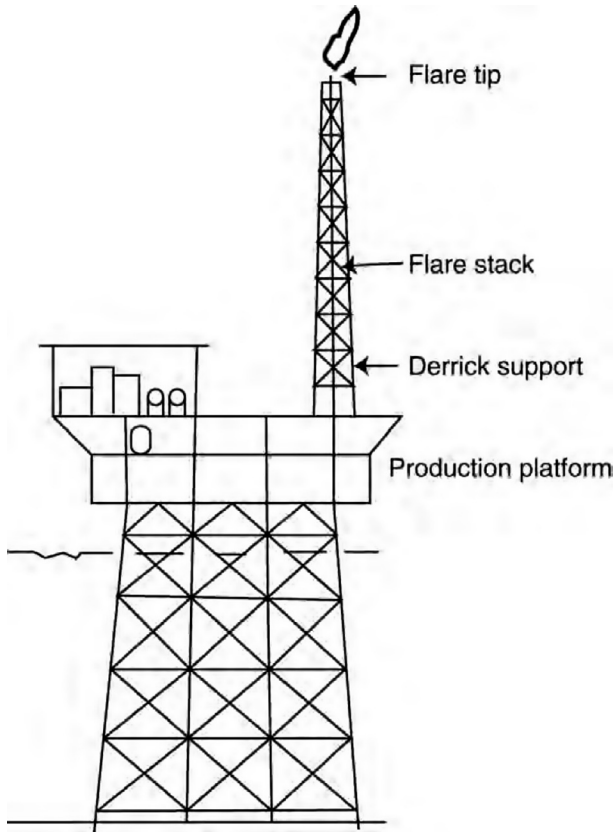


Figure 11.17 Derrick-supported flare.

11.7.10.3 Bridge-Supported Flares

In the bridge-supported flare (see [Figure 11.18](#)), the production platform is connected to a separate platform that is devoted to the flare structure. Bridges can be as much as 600 ft long, and bridge supports usually are spaced approximately every 350 ft.

11.7.10.4 Remote Flares

Remote flares (see [Figure 11.19](#)) are located on a separate platform connected to the main platform by a subsea relief line. The main disadvantage of remote flares is that any liquid carryover or subsea condensation will be trapped in pockets in the connecting line.

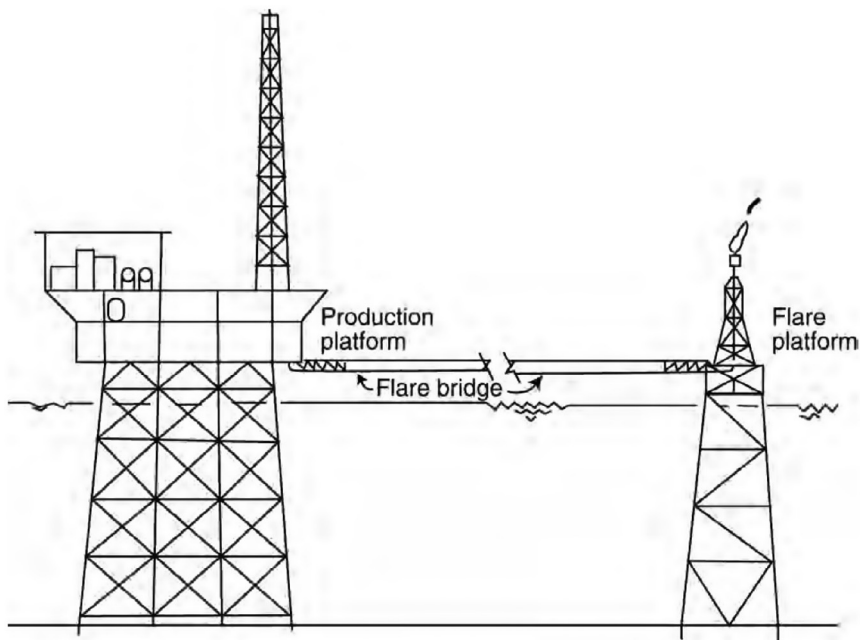


Figure 11.18 Bridge-supported flare.

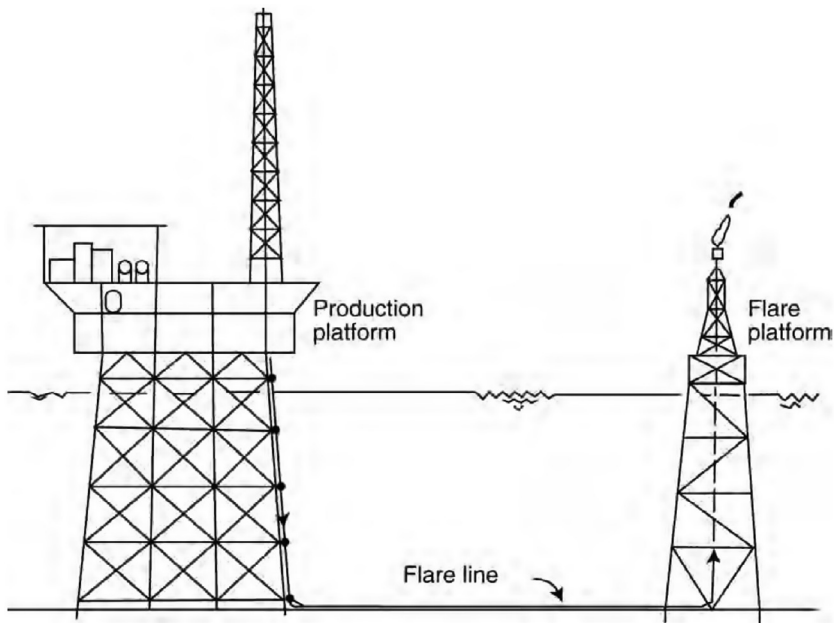


Figure 11.19 Remote flare with subsea relief line.

11.7.11 Flare-Stack Design Criteria

Important design criteria that determine the size and cost of flare stacks include flare-tip diameter and exit gas velocity, pressure-drop considerations, flare-stack height, gas dispersion limitations, flame distortion caused by lateral wind, and radiation considerations.

11.7.11.1 Flare-Tip Diameter and Exit Gas Velocity

The flare-tip diameter should provide a large enough exit velocity so that the flame lifts off the flare tip but not so large as to blowout the flare. Typically, the flare diameter and gas velocity are determined by the flare supplier. They are sized on the basis of gas velocity, although pressure drop should also be checked.

11.7.11.2 Flare-Tip Diameter

Low-pressure flare tips are sized for 0.5 Mach for a peak, short-term, infrequent flow (emergency release) and 0.2 Mach for normal conditions, where Mach equals the ratio of vapor velocity to sonic velocity in that vapor at the same temperature and pressure and is dimensionless. These API 521 recommendations are conservative [6]. Some suppliers are designing “utility-type” tips for rates up to 0.8 Mach for emergency releases. For high-pressure flare tips, most manufacturers offer “sonic” flares that are very stable and clean burning; however, they do introduce a higher backpressure into the flare system. Smokeless flares should be sized for the conditions under which they are to operate smokelessly.

11.7.11.2.1 Velocity Determination

The sonic velocity of a gas can be calculated with

$$V_s = \left(\frac{1720 k T Z}{S} \right)^{1/2} \quad (11.2)$$

Gas velocity can be determined from

$$V = \frac{(60 Q_g T Z)}{d_i^2 P_{CL}}, \quad (11.3)$$

and the critical flow pressure at the end of the relief system can be calculated with

$$P_{CL} = \frac{(2.02) Q_g}{d_i^2} \left(\frac{TS}{k(k+1)} \right)^{0.5}, \quad (11.4)$$

where d_i = pipe inside diameter, in.; k = ratio of specific heats, C_p/C_v ; P_{CL} = critical pressure at flare tip, always ≥ 14.7 , psia; Q_g = gas-flow rate, MMscf/D; S = specific gravity, ratio; T = temperature, °R; V = gas velocity, ft/s; V_s = sonic velocity, ft/s; Z = gas compressibility at standard conditions, where air = 1, psi^{-1} .

11.7.11.3 Pressure-Drop Considerations

Pressure drops as large as 2 psi have been used satisfactorily. If the tip velocity is too small, it can cause heat and corrosion damage. Furthermore, the burning of the gases becomes quite slow and the flame is influenced greatly by the wind. The low-pressure area on the downwind side of the stack may cause the burning gases to be drawn down along the stack for 10 ft or more. Under these conditions, corrosive materials in the stack gases may attack the stack metal at an accelerated rate, even though the top 8–10 ft of the flare is usually made of corrosion-resistant material.

For conventional (open-pipe) flares, an estimate of the total flare pressure drop is 1.5 velocity heads, which is based on nominal flare-tip diameter. The pressure drop is determined by

$$\Delta P_w = \frac{\rho_g V^2}{(2g)(144)} = \frac{\rho_g V^2}{9274}, \quad (11.5)$$

where g = acceleration due to gravity, 32.3 ft/s^2 ; V = gas velocity, ft/s; ΔP_w = pressure drop at the tip, inches of water; and ρ_g = density of gas, lbm/ft^3 . Figure 11.20 shows a “quick-look” nomograph to determine the flare-tip diameter.

11.7.11.4 Flare-Stack Height

The stack height is generally based on the radiant-heat intensity generated by the flame. The stack should be located so that radiation releases from both emergency and long-term releases are acceptable and so that hydrocarbon and H_2S dispersion is adequate if the flame is extinguished. The stack also should be structurally sound and withstand wind, earthquake, and other miscellaneous loadings. API RP 521, Appendix C, provides guidance on sizing a flare stack [6].

The Hajek and Ludwig equation (see API RP 521) may be used to determine the minimum distance from a flare to an object whose exposure to thermal radiation must be limited.

$$D = \left(\frac{\tau EQ}{4\pi K} \right)^{0.5}, \quad (11.6)$$

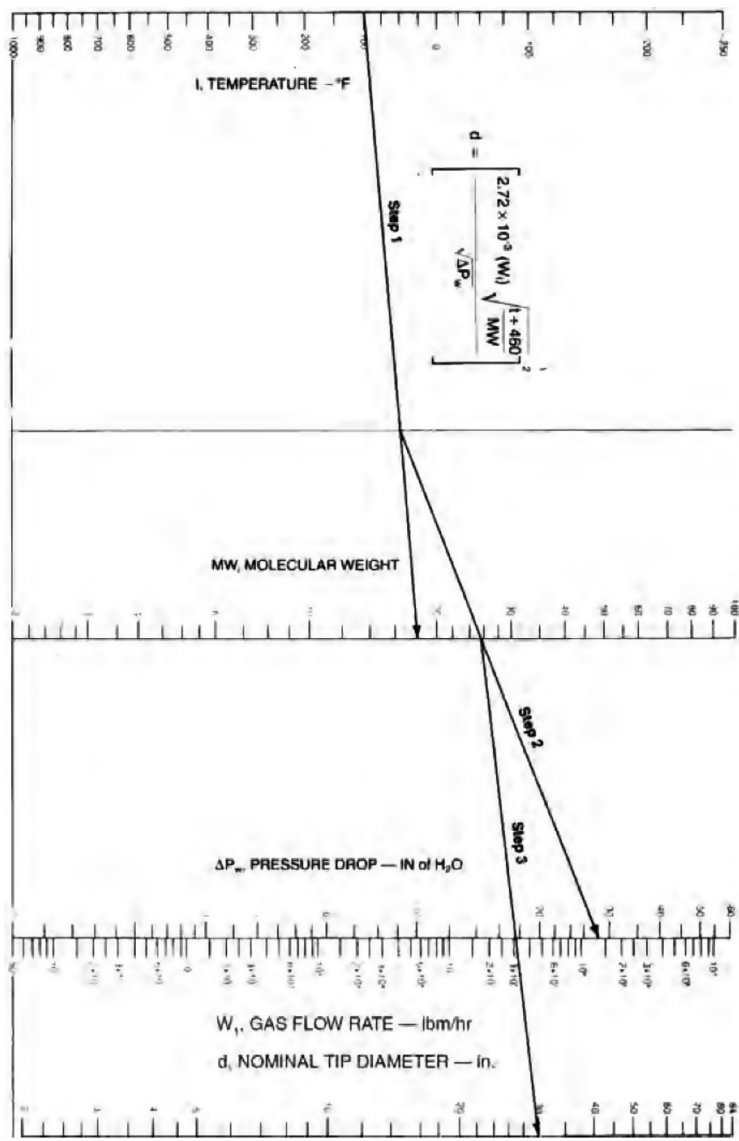


Figure 11.20 Nomograph to determine flare-tip diameter. After NAO Inc. [9].

where D =minimum distance from the midpoint of the flame to the object being considered, ft; E =fraction of heat radiated; K =allowable radiation level, BTU/h-ft²; Q =heat release (lower heating value), BTU/h; τ =fraction of heat intensity transmitted, defined by Equation (11.7).

Table 11.4 Emissivity
Component

Component	Emissivity
Methane	0.11-0.15
Butane	0.22-0.33
Natural gas	0.19-0.23

After API RP 521 [6].

Table 11.2 shows allowable radiation levels, and Table 11.4 shows component emissivity. Humidity reduces the emissivity values in Table 11.5 by a factor of τ , which is defined by

$$\tau = 0.79 \left(\frac{100}{r} \right)^{1/16} \left(\frac{100}{R} \right)^{1/16}, \quad (11.7)$$

where r = relative humidity, fraction; R = distance from flare center, ft; τ = fraction of heat transmitted, in range of 0.7-0.9.

11.7.11.5 Gas Dispersion Limitations

In some cases, it may be desirable to check the stack height on the basis of atmospheric dispersion of pollutants. Where this is required, the authorities with jurisdiction normally will have a preferred calculation method.

Table 11.5 Allowable Radiation Levels^a

Condition	Allowable Radiation (Btu/h-ft ²)		
	RP 521	One Industry Method	NAO
Areas in which operators are not likely to be. Exposure limited to a few seconds	5000	5000	-
Areas in which operators have access. Exposure limited to a few seconds	3000	1200	-
Equipment indefinitely exposed	5000	1000	3000
Areas in which emergency action lasting up to 1 min is necessary	2000	-	-
Areas in which emergency action lasting several minutes is necessary	1500	-	1500
Personnel continuously exposed	500	300	440

^aValues assume an additional solar radiation of 200-300 Btu/h-ft² is present. Normally design flare for 200-500 Btu/h-ft².

After API RP 521 [6].

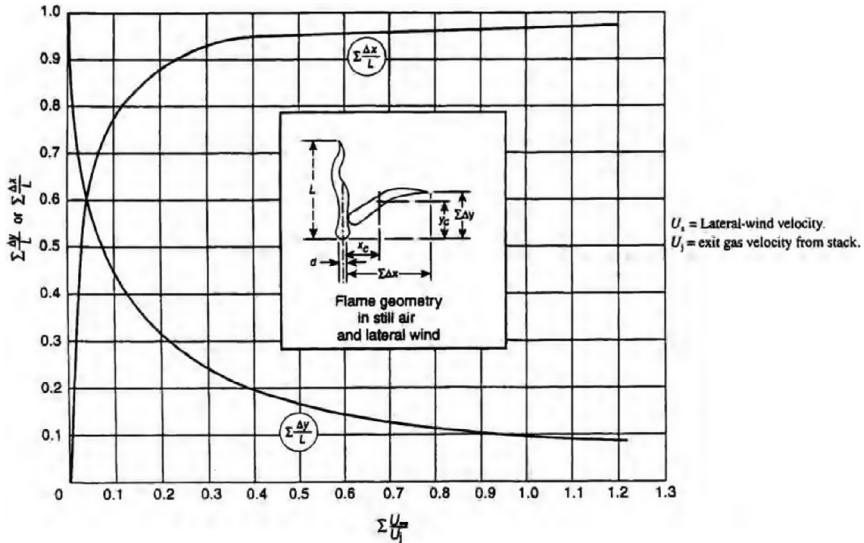


Figure 11.21 Approximate flame distortion caused by lateral wind on jet velocity from flame stack. *Reproduced courtesy of the American Petroleum Institute [6].*

11.7.11.6 Flame Distortion Caused by Lateral Wind

Another factor to be considered is the effect of wind tilting the flame, which varies the distance from the center of the flame. The center of the flame is considered to be the origin of the total radiant-heat release with respect to the plant location under consideration. [Figure 11.21](#) gives a generalized curve for approximating the effect of wind.

11.7.11.7 Radiation Considerations

There are many parameters that affect the amount of radiation given off by a flare, including the type of flare tip, whether sonic or subsonic (HP or LP) or assisted or non-assisted; emissivity of flame produced or flame length produced; amount of gas flow; heating value of gas; exit velocity of flare gas; orientation of flare tip; wind velocity; and humidity level in air.

Several design methods are used for radiation calculations. The most common methods are the API simple method and the Bruztowski and Sommers method. Both methods are listed in API *RP 521*, Appendix C [6]. These methods are reasonably accurate for simple low-pressure pipe flares (utility flare) but do not accurately model high-efficiency sonic-flare

tips, which produce short, stiff flames. The fourth edition of API *RP 521* suggests that manufacturers' proprietary calculations should be used for high-efficiency sonic-flare tips [6].

11.7.11.8 Purge Gas

Purge gas is injected into the relief header at the upstream end and at the major branches to maintain a hydrocarbon-rich atmosphere in each branch, into the off-plot relief system, and into the flare stack. The gas volume typically is enough to maintain the following velocities: ft/s for density seals, 0.4 ft/s for fluidic seals, and 0.4-3 ft/s for open-ended flares. *RP 521* states that the oxygen concentration must not be >6% at 25 ft inside the tip [6]. When there is enough PSV leakage or process venting to maintain the desired backpressure, no purge gas is injected.

11.7.12 Burn Pits

Burn pits can handle volatile liquids. They must be large enough to contain the maximum emergency flame length and must have a drain valve and pump (if required) to dispose of trapped water. The flare should be pointed down, and the pilot should be reliable. Because of the uncertainty regarding the effects of wind on the center of the flame, it is recommended that the greater of either 50 ft or 25% be added to the calculated required distance behind the tip. Burn pits should be at least 200 ft from property lines. A fence or some other positive means for keeping animals and personnel away from a potential radiation of 1200 BTU/h-ft² should be installed.

11.7.13 Vent Design

The size of a vent stack must consider radiation, velocity, and dispersion.

11.7.13.1 Radiation

The vent should be located so that radiation levels from ignition are acceptable.

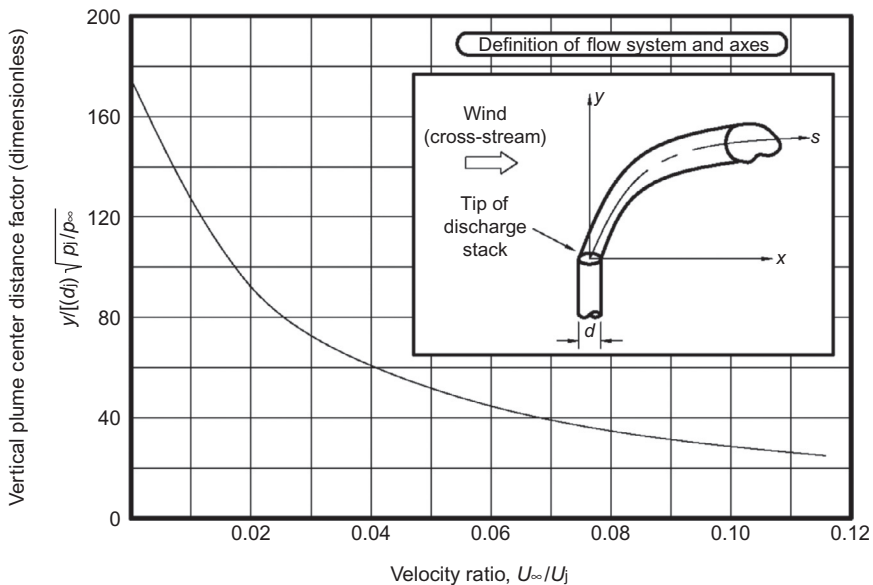
11.7.13.2 Velocity

The vent must have sufficient velocity to mix air with gas to maintain the mixed concentration below the flammable limit within the jet-dominated portion of the release. The vent should be sized for an exit velocity of at least 500 ft/s (100 ft/s minimum). Studies indicate that gases with velocities of 500 ft/s or more have sufficient energy in the jet to cause turbulent mixing with air and will disburse gas in accordance with the following equation.

$$\frac{W}{W_o} = 0.264 \left(\frac{Y}{D_i} \right), \quad (11.8)$$

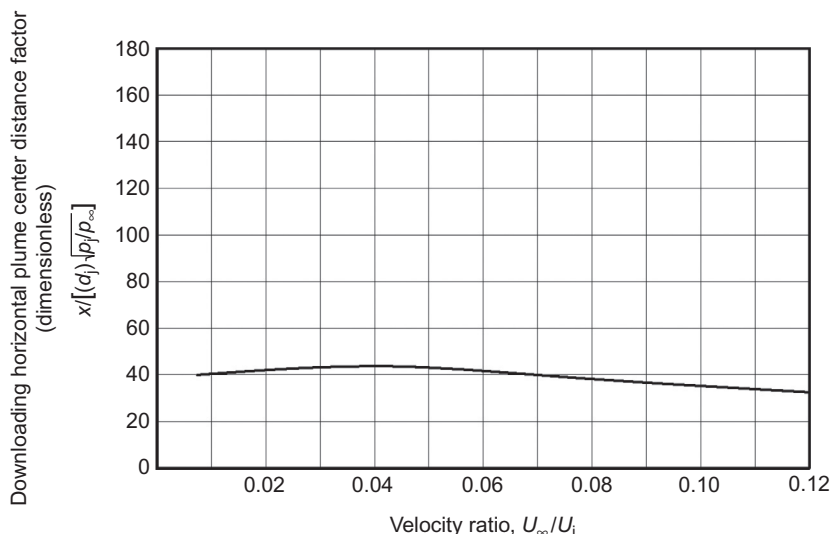
Where W = weight flow rate of the vapor/air mixture at distance Y from the end of the tailpipe; W_o = weight flow rate of the relief-device discharge, in the same units as W ; Y = distance along the tailpipe axis at which W is calculated; D_i = tailpipe diameter, in the same units as Y .

Equation (11.8) indicates that the distance Y from the exit point at which typical hydrocarbon relief streams are diluted to their lower flammable limit occurs ~ 120 diameters from the end of the discharge pipe. As long as a jet is formed, there is no fear of large clouds of flammable gases existing below the level of the stack. The distance to the lean flammability concentration limits can be determined from Figures 11.22–11.24 [6]. The horizontal limit is ~ 30 times the tailpipe diameter.



- U_∞ = wind speed, in feet per second (m/s)
- U_j = jet exit velocity, in feet per second (m/s)
- y = vertical distance, in feet (m)
- ρ_j = fluid density inside the tip exit, in pounds per cubic foot (kg/m^3)
- ρ_∞ = density of the ambient air, in pound per cubic foot (kg/m^3)
- d_j = inside diameter of the tip (jet exit diameter), in feet (m)
- a_y = distance factor multiplied by $d_j (\rho_j/\rho_\infty)^{1/2}$

Figure 11.22 Maximum downwind vertical distance from jet exit to lean flammability concentration limit (gases). *Reproduced courtesy of the American Petroleum Institute [6].*



X = horizontal distance, in feet
(Meter)

X = distance factor multiplied
by $d_i(p_i/p_\infty)^{1/2}$

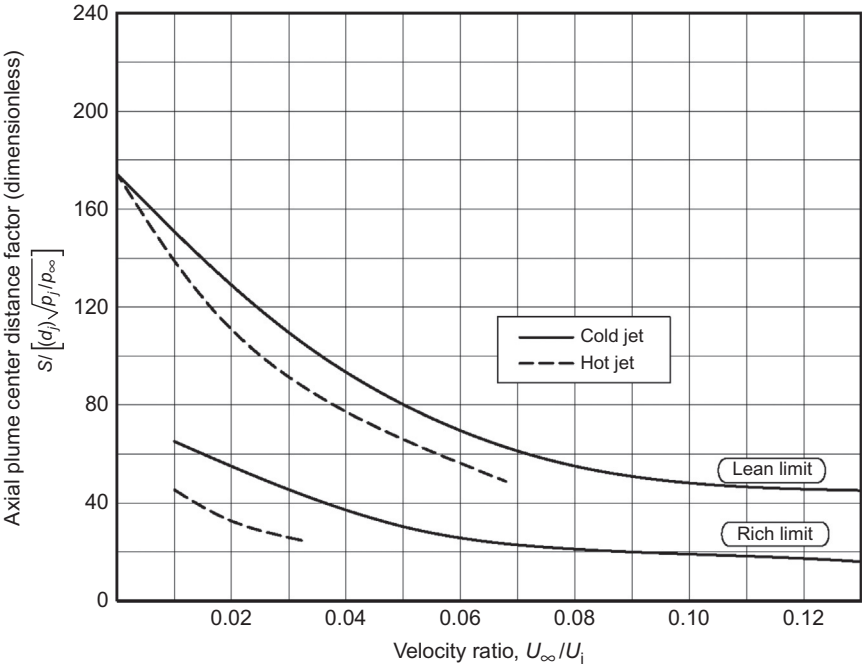
Note: See Figure 11.22 for the
nomenclature of the variables
used above and for a definition
of flow system and axes

Figure 11.23 Maximum downwind horizontal distance from jet exit to lean flammability concentration limit (Petroleum gases). *Reproduced courtesy of the American Petroleum Institute [6].*

Industry practice is to locate vent stacks 50 ft horizontally from any structure running to a higher elevation than the discharge point. The stacks must vent at least 10 ft above any equipment or structure within 25–50 ft above a potential ignition source. Because the flame can be ignited, the height of the stack must be designed or the pit located so that the radiation levels do not violate emergency conditions.

11.7.13.3 Dispersion

The vent must be located so that dispersion is adequate to avoid potential ignition sources. The dispersion calculation of low-velocity vents is much more difficult and should be modeled by experts familiar with the latest computer programs. Location of these vents is very critical if the gas contains H_2S because even low concentrations at levels accessible by personnel could be hazardous. The location of low-velocity vents should be checked for radiation in the event of accidental ignition.



S = horizontal distance, in feet
(m)
S = distance factor multiplied
by $d_i (p_i / p_\infty)^{1/2}$

Note: See Figure 11.24 for the
nomenclature of the variables
used above and for a definition
of flow system and axes

Figure 11.24 Axial distance to lean and rich flammability concentration limits (petroleum gases). *Reproduced courtesy of the American Petroleum Institute [6].*

SI Metric Conversion Factors

Btu	×	1.055 056	E+00	=kJ
Btu/hr	×	2.930 711	E+01	=W
ft	×	3.048*	E-01	=m
ft/s	×	3.048*	E-01	=m/s
ft/s ²	×	3.048*	E-01	=m/s ²
ft ²	×	9.290 304	E-02	=m ²
ft ³	×	2.831 685	E-02	=m ³
°F		(°F-32)/1.8		=°C
lbm	×	4.535 924	E-01	=kg
lbm/ft ³	×	1.601 846	E+01	=kg/m ³
psi	×	6.894 757	E+00	=kPa
°R			°R/1.8	=°K

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Case Study: Membrane/Amine Hybrid Grissik Gas Plant [1–3], Sumatra, Indonesia



A.1 INTRODUCTION

ConocoPhillips operates the Grissik Gas Plant ([Figure A.1](#)) on behalf of its partners:

- Talisman Energy
- Pertamina
- BPMigas

Design basis:

- Gas feed: 310 MMscfd
- CO₂ concentrations:
 - Inlet: 30%
 - Outlet: 3%



A.2 PROCESS OVERVIEW

The CO₂ removal process uses a membrane/adsorption hybrid process.

The hybrid process relies on:

- Membrane separation
- Amine adsorption

A simplified process flow diagram is shown in [Figure A.2](#).

The thermal swing adsorption (TSA) unit:

- Removes heavy hydrocarbons
- Serves three functions
 - Membrane
 - Pretreatment
 - Feed gas dehydration

- Sales gas hydrocarbon dew pointing

Benefits of the membrane/amine hybrid process:



Figure A.1 Grissik gas plant.

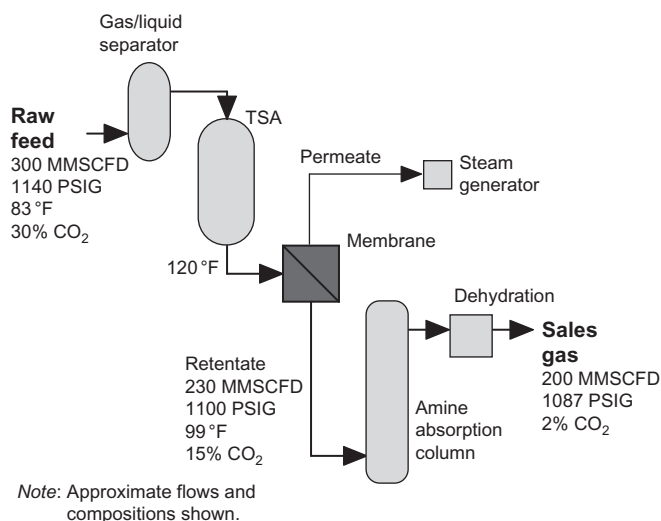


Figure A.2 Grissik process flow diagram.

The process has a single-stage membrane and utilizes the thermal value in the permeate stream.

By employing this kind of single-stage membrane, the process benefits from the simplicity of a membrane separation process without the use of a recycle compressor.

The process avoids hydrocarbon losses.

The process sends CO₂-rich permeate to an atmospheric burner to produce steam that is used in the amine plant for regeneration.

Natural gas exiting the membrane:

Contains about 15% CO₂.

Is fed to the amine absorption column where CO₂ is removed to about 3%.

Permeate rich in CO₂ exits the membrane at near atmospheric pressure.



A.3 BACKGROUND

A.3.1 General Considerations

The plant was commissioned and built in 1998 without TSA membrane pretreatment. Initial well tests indicated minimal amounts of heavy hydrocarbons. Subsequent tests showed that this was not the case, however.



A.4 FIRST COMMISSIONING

The membrane initially installed with pretreatment consisted of the following components:

- Coalescing filter

- Nonregenerable absorption guard bed

At startup in 1998:

- The levels of heavy hydrocarbons (CO_{10+} , aromatics, and naphthenes) were higher than anticipated.

- The higher levels of heavy hydrocarbons resulted in a sharp reduction in membrane capacity (declining to 20–30% of initial capacity within in a month).

- To maintain production capacity, the membrane elements were frequently replaced.

Installation of the TSA unit:

- ConocoPhillips evaluated the heavy hydrocarbon removal processes including the following:

 - Gas chilling process

 - Regenerable adsorption process

Gas chilling process:

- The process was deemed ineffective at the plant operating pressure, which was near the cricondenbar of the feed gas phase envelope.

Regenerable adsorption process:

- The regenerable adsorption process is a short-cycle process from Engelhard that uses Sorbead (Silica Gel) as the adsorbent.

The process uses multiple beds in parallel adsorption to remove:

- Heavy hydrocarbons

- Aromatics

- Naphthenes



Figure A.3 Engelhard thermal swing adsorption unit.

The adsorption cycle is followed by the regeneration of the silica gel at elevated temperatures.

The TSA unit was built and installed by Kvaerner in 2000.

The TSA unit was designed to reduce C_{6+} components (including aromatics and naphthenes) so that membrane performance could be maintained for an extended period of time.

The unit was designed with two separate trains, each with four adsorption vessels (refer to [Figure A.3](#)).

A.5 TSA DESIGN AND PERFORMANCE

A.5.1 General Design Considerations

Given that the feed gas was found to contain high levels of heavy hydrocarbons (C_{10+} , aromatics, and naphthenes), the TSA unit had two functions and solved two problems:

The TSA unit removed heavy hydrocarbons for proper pretreatment so as to yield long membrane life.

The TSA unit removed heavy hydrocarbons, allowing the sales gas to meet hydrocarbon dew point specs.

Given that water adheres more strongly to the Sorbead adsorbent than to any of the hydrocarbons, the TSA unit also dehydrated the feed upstream of the membrane unit.

A.6 TSA PROCESS DESCRIPTION

Each train was designed to treat 225 MMscfd.

Each train consists of four internally insulated adsorber towers that allow the system to:

Minimize the thermal mass for the short thermal cycle.

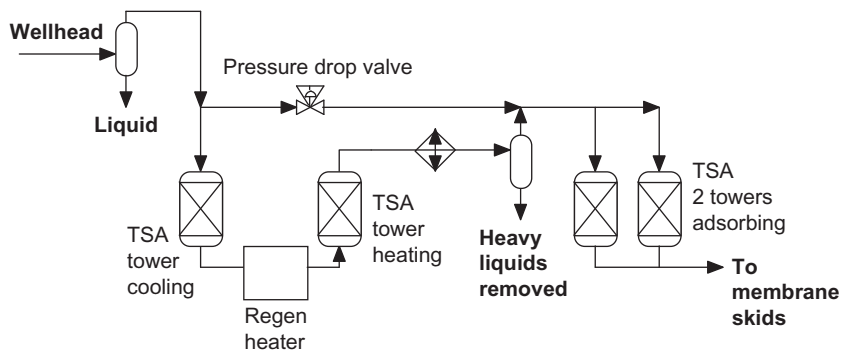


Figure A.4 TSA process flow diagram.

Reduce the heat load on the system.

The TSA process flow diagram (Figure A.4) illustrates the following steps: After passing through the two-phase separator, the feed gas is split into two parallel paths.

The majority of the gas flows through the pressure drop valve and then directly to two parallel adsorption towers.

The cycle time of the towers is staggered by 50% to allow for a continuous flow of treated gas to the downstream membrane unit.

The balance of the feed gas bypasses the pressure drop valve so as to provide the necessary flow through the towers being cooled and heated.

The regeneration path contains the following components:

- Tower being cooled
- Regeneration heater
- Tower being heated
- Heat recovery heat exchangers
- Spent regeneration gas separator

Each tower is associated with six valves that allow it to change position for the following functions:

- Adsorbing
- Heating/regenerating
- Cooling
- Adsorbing

Wet feed gas is used as the regeneration medium, and because of the pressure drop valve, there is no need for a compressor to boost the pressure of the spent regeneration gas.

During adsorption:

Water and C_{6+} components are adsorbed at 1100 psig and 90–140 °F.

Prior to C_{6+} breakthrough, the tower position is switched to heating mode and is completely heated to 540 °F.

Internal insulation allows for the heating of the adsorbent only and not the steel shell.

During heating, the water vapor and C_{6+} components are desorbed. The spent regeneration gas stream containing water and C_{6+} is then cooled, resulting in the following:

Condensed liquids are removed in the regeneration gas separator.

The heavy hydrocarbons exit the system (only at this stage).



A.7 REASONS FOR FOUR TOWERS

In order to maintain an acceptable flow velocity across the adsorber bed, the number of towers used is a function of two factors:

Flow rate

Tower diameter

The maximum tower diameter was determined by transport limits, and the Grissik design resulted in four towers with two towers engaged in parallel adsorption. This design offers the following features:

Internal insulation minimizes the amount of regeneration gas required.

Heating and cooling towers are in a series, which leads to a reduction in the amount of regeneration gas required.

Parallel adsorption towers provide an equalized composition of the treated gas.

In a single-tower system, gas composition differs between the beginning and end cycles because of the breakout of individual components. In a four-tower system, on the other hand, two towers engaged in adsorption provide the following benefits:

There is an offset time of half an adsorption cycle.

The gas composition of the combined outlet gas is more constant than the composition of outlet gas from a single-tower system.



A.8 CYCLE TIMES AND BREAKTHROUGH

Cycle times demonstrate the following features:

Cycle times are driven by the breakthrough behavior of the C_{6+} components in the given tower design, as they relate to the hydrocarbon specifications for the treated gas.

Table A.1 Tower Mode Timing

Tower 1	2 h adsorption	1 h heating	1 h cooling
Tower 2	1 h cooling	2 h adsorption	1 h heating
Tower 3	1 h heating	1 h cooling	2 h adsorption
Tower 4	1 h adsorption	1 h heating	1 h cooling, 1 h adsorption

Cycle times result from analysis and field observations.

A typical cycle consists of the following stages (Refer to [Table A.1](#)):

Two hours of adsorbing

One hour of heating

One hour of cooling



A.9 HEAT RECOVERY BETWEEN COOLING AND HEATING

The system uses one tower for heating and one tower for cooling at a time.

As it leaves the tower being cooled, hot gas flows through the heater in order to get additional heat in.

At the beginning of the cycle, gas exiting the tower in the process of cooling is almost at the required heating temperature.

This exchange of gas results in almost no make-up heat being required

Due to the entire tower being cooled, the gas is at the hot regeneration temperature of 540 °F.

During the cooling cycle, the temperature of the gas exiting the cooling tower drops, so that the heater has to provide the required heating gas temperature.

Gas-to-gas heat exchanger:

The gas-to-gas heat exchanger is used to capture the heat exiting the tower being heated.

Hot gas is cross-exchanged with the gas upstream of the regeneration gas heater (refer to [Figure A.4](#)).

The exchanger is bypassed during the time when the gas exiting the cooling tower is at a higher temperature than the gas leaving the tower in the heating step.

Regeneration heater:

The regeneration heater employs direct-fired heating elements.

Table A.2 TSA Hydrocarbon Dew Point

TSA feed gas	86° Fat 1150 psig
TSA outlet gas	−22° Fat 1115 psig

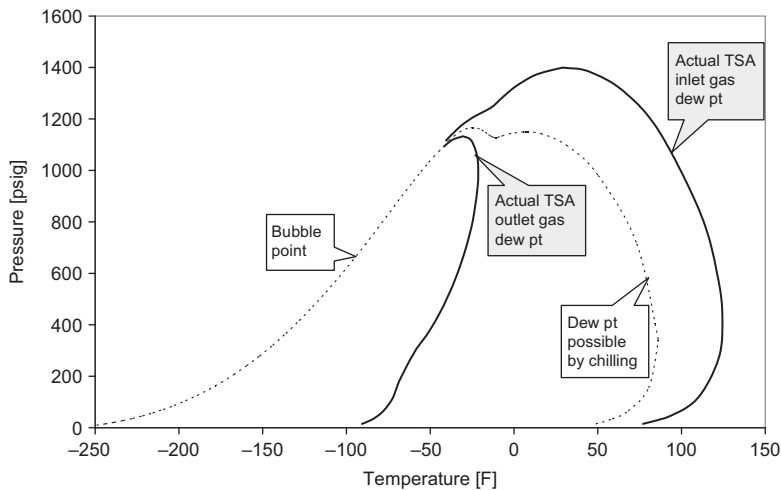


Figure A.5 TSA inlet and outlet phase envelopes.

The size of the heater depends on the regeneration gas flow required to heat the adsorption bed and desorb the water and hydrocarbons within the design cycle time.

TSA performance after the plant was recommissioned in October 2000:

The TSA system’s improved removal of heavy hydrocarbons led to excellent membrane performance.

The TSA system’s performance related to the hydrocarbon dew point was impressive (see [Table A.2](#)).

Corresponding phase envelopes are shown in [Figure A.5](#).

[Figure A.6](#) shows the results of gas-sampling conducted with a mass spectrometer where both the feed and exit streams of the TSA system were analyzed dynamically. The ratio of hydrocarbon concentration in the outlet versus the inlet is shown.

Note the strong cutoff that occurs between C₆ and C₈.

Heavier hydrocarbons are essentially completely removed.

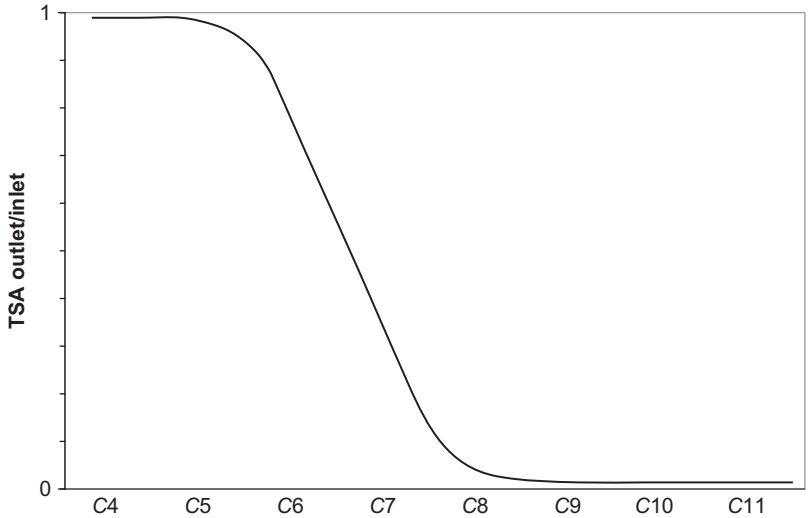


Figure A.6 TSA hydrocarbon tail.

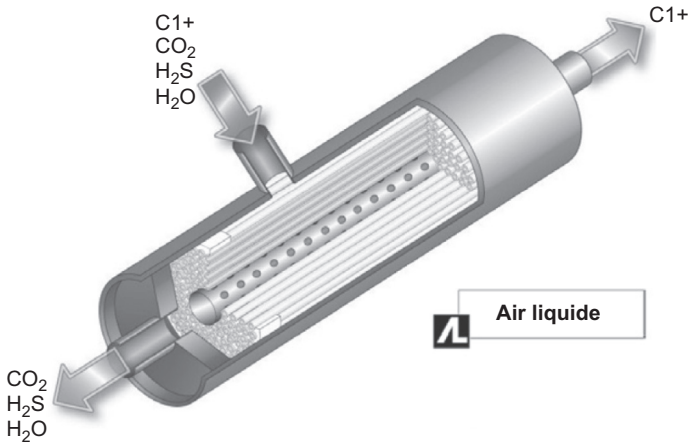


Figure A.7 Air Liquide Medal natural gas membrane.



A.10 AIR LIQUIDE MEDAL MEMBRANE

A.10.1 General Considerations

Polyimide hollow fiber membrane elements (shown in Figure A.7) provide for the high efficiency separation of CO₂ from hydrocarbon streams.



Figure A.8 Skid containing Air Liquide membrane elements.

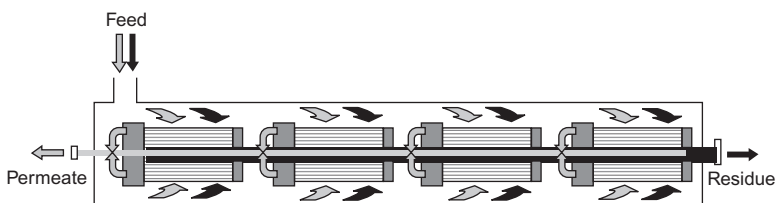


Figure A.9 Multiple membrane element flow arrangement.

The membrane system was fabricated as multiple skids (refer to [Figure A.8](#)) operating in parallel.

Each skid contains multiple horizontal tubes.

Each tube contains multiple membrane elements (refer to [Figure A.9](#)) with the following features:

Multiple elements are installed in a single tube.

Membrane elements actually function in parallel.

More than 100 membrane elements are used in this plant.

Feed gas enters the tube near one end and flows axially to all the membrane elements by way of an annular clearance.

Each element is composed of several hundred thousand parallel hollow polyimide fibers.

Feed gas enters the membrane elements on the interior fiber shell and flows over the fibers, where CO_2 is removed, to a coaxial tube in the center of each element (retentate).

Retentate streams for each element flow axially to exit at one end of the tube.

CO_2 selectively permeates into the bore of the fibers and then flows axially to a collection point at the end of each element (permeate).

The permeate of each element is then collected in the coaxial center tube and flows axially to exit the tube at the end opposite the retentate.

A.11 MEMBRANE PERFORMANCE

Typical operating conditions:

Membrane skids are fed directly from the output of the TSA.

Feed temperatures vary between 90 and 120 °F.

Feed pressure is 1100 psig.

Feed gas contains 30% CO₂.

The permeate pressure is about 10 psig, and the gas flows to the steam generator burners.

Hydrocarbon losses versus time:

One of the major advantages of the polyimide membrane is its ability to maintain integrity indefinitely, even when aging in the presence of heavy hydrocarbons.

As shown in Figure A.10, membrane integrity is solid, and the hydrocarbon losses have decreased somewhat since startup.

This trend of decreasing hydrocarbon losses indicates no loss of membrane integrity and actually shows a slight increase in apparent intrinsic membrane selectivity.

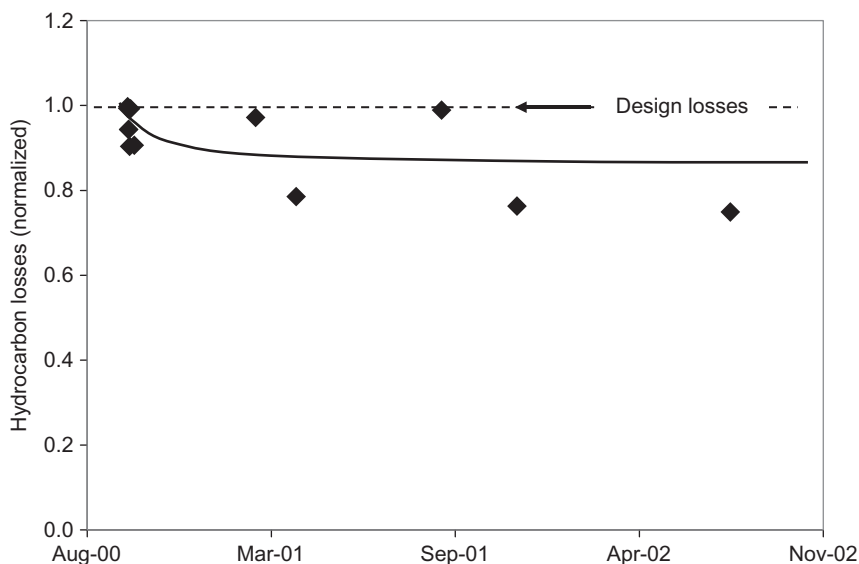


Figure A.10 Membrane hydrocarbon losses versus time.

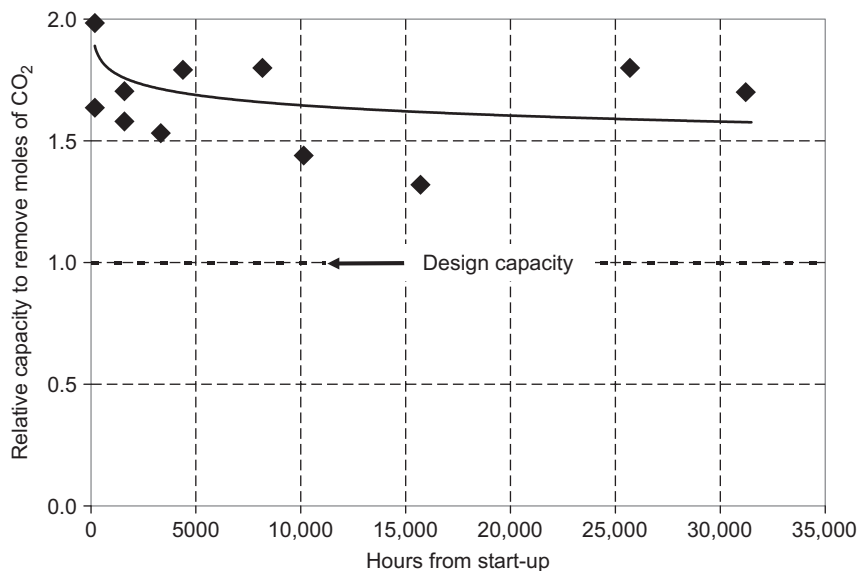


Figure A.11 Membrane capacity versus time.

Such a selectivity increase would be consistent with a change in permeability (see below).

Membrane capacity versus time after the TSA system was commissioned in October 2000:

One membrane skid was retrofitted with new membrane elements and its performance tracked.

The results are shown in [Figure A.11](#).

The vertical axis is labeled “relative capacity to remove moles of CO₂,” which is the normalized intrinsic membrane permeability.

The initial capacity was well above expectations based on the design, and after 10 years of operation, the capacity still remains above the design expectations.

The exact membrane life can be extrapolated to be over 12 years without replacement.

The excellent operation of the TSA system and membrane have resulted in years of trouble free operation with zero maintenance, that is, no membrane replacements.

The membrane skids were shut down and restarted many times to allow for the maintenance of surrounding equipment or capacity turndown. Start and stop or pressurization and depressurization cycles have no effect on membrane performance, although caution must be used to avoid reverse pressurization.



A.12 PERMEATE/ACID GAS UTILIZATION

Two waste heat boiler units are installed. These units have the following features:

Waste heat boilers recover waste heat available in the low-BTU permeate gas stream (150–250 Btu/scf) from the membrane units.

Utilizing the waste heat from the permeate stream means a single-stage membrane can be used without the limitations of a second membrane stage with the accompanying recycle gas compressor, while still avoiding hydrocarbon losses.

Boilers are designed to incinerate the acid gases removed by the amine unit.

Auxiliary fuel is utilized to make up any inadequacy of heating value input and to stabilize the flame.

The furnace temperature is maintained above 1600 °F prior to introducing permeate fuel or acid gases.

Lower temperatures lead to the incomplete destruction of the component and result in emission hazards.

Waste heat boilers are controlled by a steam header that actuates pressure control valves on each steam drum.

The output of the steam header pressure controller goes through flow ratio controllers for permeate gas, fuel gas, and combustion air.

The fuel gas flow rate is set at around 10% of the permeate gas flow rate, while combustion air is controlled to ensure stoichiometry and complete combustion with 2–5% excess air.

A waste heat boiler produces steam up to 210,000 lbs/h at 150 psig and 348 °F.

The biggest consumer of the steam produced is the amine system.

Condensing heat released by the steam is used to remove acid gas from the amine solvent in the amine reboilers.



A.13 AMINE SYSTEM

The amine system further reduces CO₂ and H₂S to meet sales gas specifications.

The residue gas from the membrane unit, containing 15% CO₂, flows into the amine contactors and contacts the lean amine (50 wt%-activated Methyl-diethanolamine (MDEA)).

CO₂ absorption by activated MDEA is limited to a maximum loading of 0.5 mol acid gas/mol MDEA.

The CO₂ content in the treated gas varies between 2% and 5% by volume (3 vol% average).

Rich amine is then flashed at 75 psig, heated through a lean/rich amine exchanger, and regenerated by the steam-heated reboiler.

The 150-psig steam used for regenerating amine is produced in the waste-heat boiler that burns permeate gas.

Several common problems of an amine system include the following:

- Reduced strength and ability to absorb acid gas

- Degradation

- Foaming

- CO₂ corrosion attack during acid gas breakout inside the reboiler

Most problems found in an amine system result from the presence of contaminants in the amine solvent, including the following:

- Heat-stable salts

- Degradation products

- Injected chemicals

- Hydrocarbons

- Particulates

Heat-stable salts and degradation products are formed by amine solvents that decompose and/or react with other contaminants.

The TSA system and membrane installed upstream of the amine system has mitigated the above problems to an acceptable level as follows:

The TSA unit removes heavy hydrocarbons from the feed gas, and it has nearly eliminated foaming risk in the amine solvent.

An antifoam injection system is provided to anticipate worst-case conditions.

CO₂ content reduction by the membrane unit also leads to the following occurrences:

- Breakout develops in the regeneration process

- CO₂ breakout in the regeneration process is reduced.

- Contaminants that may trigger salt formation or amine degradation decrease in volume. (Contaminants could also be introduced by makeup water or even makeup amine, however.)

The condition of the treated gas at the outlet of the amine unit is normally 3 vol% CO₂ and 2–4 ppmv H₂S, while the sales gas contract specifies 5 vol% CO₂ and 8 ppmv H₂S.

One advantage of high performance absorption is its potential to increase system deliverability by bypassing some untreated gas and blending with treated gas, while maintaining the sales gas specification.

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Case Study: The Judge Digby Gas Plant Hikes Production with Quick Solvent Change-Out [1]

B.1 JUDGE DIGBY PLANT

The gas plant is operated by BP and is located in Pointe Coupee Parish in South Louisiana.

The plant came online in 1970.

Table B.1 provides the inlet feed stream composition.

The plant's Diethanolamine (DEA) amine unit produces a sales gas with <3% CO₂ and 8 ppm H₂S.

The unit consists of two trains with the following features:

Train 1 consists of a 150-MMcfd conventional amine unit using 30% DEA followed by a TEG dehydration unit.

Train 2 consists of a 100-MMcfd TEG dehydration unit only.

Figure B.1 provides an aerial view of the plant, with Train 1 shown in the upper left-hand corner.

Figure B.2 is the process flow diagram of Train 1, as described below:

The plant was initially designed to produce 150 MMcfd of treated gas. Train 1 was designed to remove 95% of the acid gas from 120 MMcfd of raw gas in the sweetening unit.

A 30-MMcfd stream of raw gas bypasses the amine unit and combines with amine plant-treated gas to yield 150 MMcfd that meets the desired CO₂ and H₂S specifications.

The bypass gas is treated with a nonrecoverable H₂S-scavenging chemical before being blended with the sales gas.

In 1999, an engineering study assessed the unit's performance as follows:

The maximum throughput was at a capacity of 135–140 MMcfd and declined to 135 MMcfd.

The unit became unstable and required 24-h manned operation at production rates >140 MMcfd.

The DEA system also exhibited several problems:

The DEA solution had degraded, and the reboilers were severely fouled.

Component	Mole Percent
CO ₂	8.08
C ₁	89.54
C ₂	1.15
C ₃	0.18
I-C ₄	0.08
n-C ₄	0.05
1-C ₅	0.05
n-C ₅	0.03
C ₆₊	0.57
N ₂	0.27
H ₂ S ppm	40



Figure B.1 Judge Digby gas-processing plant.

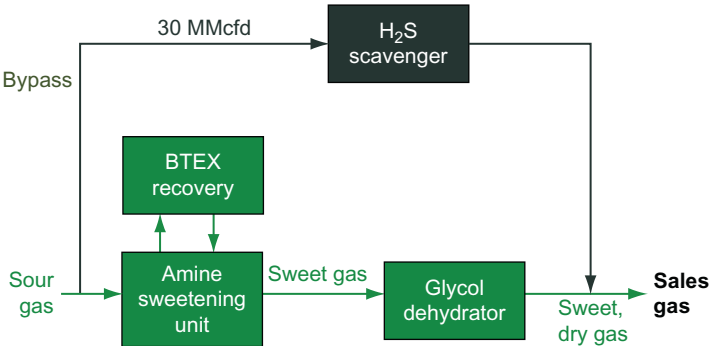


Figure B.2 Judge Digby flow diagram.

Table B.2 Operating History with DEA

	August 1999	July 2000
Inlet gas flow rate (MMcfd)	92.5	135.0
Bypass gas flow rate (MMcfd)	0	13.5
Gas flow rate to absorber (MMcfd)	92.5	122.5
Inlet gas pressure (psig)	994	1000
Inlet gas temperature (°F)	96	95
DEA circulation rate (gpm)	680	953
DEA concentration (wt%)	30	33
Lean DEA loading (mol CO ₂ /mol DEA)	0.02	0.06
Lean DEA temperature (°F)	113	110
Reboiler duty (MMbtu/h)	39.5	50
CO ₂ in outlet gas (%)	2.72	2.75

Corrosion probes indicated a high degree of corrosion.

The regenerator still was hydraulically limited and unable to fully strip the rich DEA solution.

Lean solution loadings of 0.6–0.7 mol of CO₂/mol of DEA were typical.

The fully stripped DEA should contain <0.02 mol of CO₂/mol of DEA.

High solution lean loadings and solution degradation products usually lead to corrosion and reboiler fouling problems.

High CO₂-loading in the DEA from the regenerator also prevented the absorber from removing all the CO₂ from the raw gas; therefore, there was too much CO₂ in the gas leaving the DEA unit.

Instead of treating and bypassing 122.5 and 30 MMcfd, respectively, the plant could only treat and bypass 122.5 and 13.5 MMcfd, respectively, to produce a combined sales gas volume of only 135 MMcfd that contained <3% CO₂.

Table B.2 shows the plants' performance using DEA.



B.2 DEBOTTLENECKING

BP considered several options to regain lost capacity.

The contactor was hydraulically limited to a feed rate of 120 MMcfd, which also limited BP's options. Flow rates greater than the contactor's hydraulic limit resulted in the following:

- Large amine losses
- Increased corrosion
- Operating instabilities

To maximize production, BP wanted to maximize:

- Bypass gas flow rate

- Amount of CO₂ removed in the absorber

BP also wanted to reduce the number of plant upsets at high flow rates for the following reasons:

- The plant was unmanned 16 h/day.

- Upsets during unmanned periods increased the number and cost of off-shift operator callouts.

- Each unplanned shutdown had adverse effects on producing wells.

BP decided to replace the existing DEA chemical solvent with Dow's AP-814 solvent for the following reasons:

- AP-814 requires less regeneration duty.

- AP-814 absorbs more CO₂.

Solvent change-out required the plant to take the following limited measures:

- Less than 24 h without extensive system cleanout

- No mechanical equipment modifications



B.3 PREPARING FOR THE CONVERSION

Figure B.3 shows the following components:

- Amine contactor (background)

- Regeneration still

- Benzene-toluene-ethylbenzene-xylene (BTEX) stripper (foreground).

Figure B.4 shows a process flow diagram of the Train 1 amine system.



Figure B.3 Amine sweetener including a contactor, regenerator, and BTEX stripper.

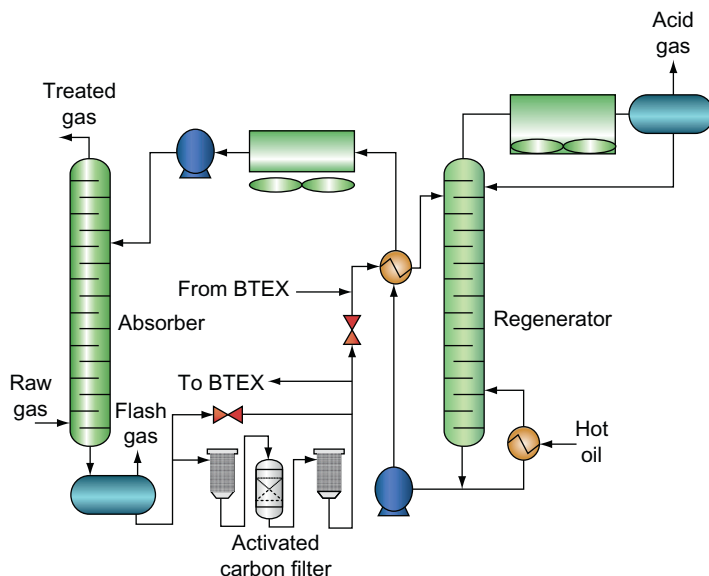


Figure B.4 Amine sweetening unit.

BP performed a gamma scan of the absorber and regenerator before the solvent switch.

An engineering analysis of the reboiler showed that it was severely fouled. Two cleaning crews manned the shutdown to ensure timely cleaning of the reboiler tubes. The cleaning process involved the following steps:

The cleaning process was the time-limiting step.

32 h were spent cleaning the severely fouled tubes before halting the process.

DEA was removed from the storage tank and shipped to another plant.

The storage tank was cleaned and filled with an initial charge of 45 wt% AP-814.

Two antifoams were brought to the plant for the following reasons:

One antifoam was used for start-up that would prevent foaming created by solids.

One antifoam was used for normal operations.

The initial plant configuration did not allow for the continuous addition of antifoam to the stripper; thus, an antifoam charge pump was added.

Process safety management considerations:

A management of change document was developed.

A process hazards analysis review was performed.

The crew considered all the changes needed to accommodate a solvent switch, including special issues such as:

- Metallurgy
- Equipment configuration
- Pump design
- Gasket materials

The review indicated that no additional modifications were required.

Training conducted for unit operators and company engineers focused on:

- New laboratory methods
- Operating techniques
- Operating parameters with the new AP-814 solvent

Operations designed a time window for the reboiler cleaning and solvent change-out.



B.4 THE TURNAROUND

In October 2000:

DEA was removed from the sweetening unit and drained from all the low points as much as possible.

The amine unit was flushed with water drained.

The activated carbon filter was emptied and filled with a fresh charge.

After start-up, a gas chromatographic analysis indicated that only trace amounts of DEA remained in the system.

Solvent swap:

Initially, inlet and outlet isolation valves were installed to isolate the reboilers.

The isolation valves leaked, and the reboilers had to be cleaned before fresh solvent was added to the unit.

The solvent swap would have taken 4 h if the isolation valves did not leak.

Reboiler tube fouling:

The reboiler tubes were hydroblasted in an attempt to remove the iron carbonate fouling.

After 32 h hydroblasting was stopped, and the reboiler was put into service.

20–30% of the tubes remained plugged with iron carbonate.

After the solvent swap and the hydroblasting of the tubes, the unit started up smoothly with the new solvent, and the plant produced the expected additional 20 MMcf/d of gas, even with the fouled reboiler tubes.



B.5 PLANT OPERATIONS

The plant operated for 6 weeks before well production was lost.

Table B.3 shows performance data collected during that time.

Given that the new solvent removed more CO₂, sweetened gas could be mixed with untreated gas.

The new solvent allowed for more operating flexibility, which helped compensate for the less-than-optimum reboiler performance.

Damaged wells were restarted in March 2001, with the following results:

The plant could not run at high rates because the reboiler tube condition had worsened.

The reboiler tube bundle was replaced.

Table B.4 shows the plant performance with AP-814 and the historical maximum performance with DEA.

After bundle replacement:

The plant had no problem meeting design capacity.

At \$3.00 Mcf⁻¹, the improved production translated to an incremental increase in sales revenues of approximately \$34 million/year.

Plant performance:

OPEX remained the same with the AP-814 despite the capacity increase.

The amine unit could theoretically process up to 168 MMcfd if other bottlenecks were removed.

Lab tests did not show any solvent degradation.

Solvent losses were low.

The sweetening unit ran for months without any significant change in solvent concentration.

Table B.3 Operating Performance with Ucarsol

Inlet gas flow rate (MMcfd)	131
Bypass gas flow rate (MMcfd)	26
Gas flow rate to absorber (MMcfd)	105
Inlet gas pressure (psig)	1001
Inlet gas temperature (°F)	102
Solvent circulation rate (gpm)	840
Lean solvent loading (mol CO ₂ /mol solvent)	0.03
Lean solvent temperature (°F)	117
Reboiler duty (MMbtu/h)	48
CO ₂ in outlet gas (%)	0.01

Table B.4 Optimized Performance Comparison

	DEA (33 wt%)	Ucarsol (45 wt%)
Maximum processing capacity (MMcfd)	136	150
Solvent circulation rate (gpm)	953	1000
Solvent CO ₂ loading (mol CO ₂ /mol solvent)		
Lean	0.06	0.03
Rich	0.49	0.46
Reboiler duty (MMbtu/h)	50	50
CO ₂ in treated gas		
Outlet from contactor (%)	2	0.2–0.5
Sales gas with bypass (%)	3	<2.5

During the change-out, the 10- μ m particulate filters were replaced with 5- μ m filters.

The new 5- μ m filters are changed-out half as often.

Corrosion is low.

Corrosion monitoring probes that indicated high levels of corrosion with DEA are now reading low levels.

Low iron levels in the solvent and good performance from the 5-micrometers filters confirm this finding.

The sweetening unit performed well during upstream upsets and well outages.



B.6 BTEX EMISSIONS

The Judge Digby gas stream contains some BTEX.

BTEX compounds have a high fuel value.

BTEX compounds are regulated for human contact and air emissions.

BTEX is more soluble in amine solvents than other hydrocarbons.

The plant has a BTEX-removal unit that strips BTEX from the rich amine, and the unit has the following features:

The unit is on the rich-amine line between the rich-amine filters and lean-rich heat exchanger.

Fuel gas is the stripping agent.

This gas, along with any gas stripped out with the BTEX, is recovered and routed to the plant's fuel-gas system.

The BTEX unit was evaluated using AP-814 solvent at a higher circulation rate. The evaluation indicated the change from DEA did not significantly alter the load on the BTEX-removal system.

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