GPSA ENGINEERING DATABOOK ERRATA
(2004 SI Edition)

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7-1	Figure 7-1, Change figure reference
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SECTION 3

Measurement

The information presented in this section provides sufficient information for determining flow quantities with a reasonable degree of accuracy, but not necessarily to the accuracy desired for custody transfer. Agreement of acceptable accuracy for custody transfer should be between the parties involved, and supplemental information and procedures may be required, such as the API Manual of Petroleum Measurement Standards or corresponding ISO standards.

FIG. 3-1

Nomenclature

C = Pitot tube flow coefficient
C' = the product of multiplying all orifice correction factors
CNT = volume indicated by the number of pulses or counts
C_{pl} = liquid pressure correction factor. Correction for the change in volume resulting from application of pressure. Proportional to the liquid compressibility factor, which depends upon both relative density and temperature.
C_{ps} = correction factor for effect of pressure on steel. See API Manual of Petroleum Measurement Standards, Chapter 12, Section 2
C_g = gravity correction factor for orifice well tester to change from a gas relative density of 0.6
C_{tl} = liquid temperature correction factor. Proportional to the thermal coefficient which varies with density and temperature
C_{ts} = correction factor for effect of temperature on steel
C_u = velocity of sound in the gas non-flowing condition.
d = orifice diameter, mm
D = internal pipe diameter of orifice meter run or prover section, mm
DL = Minimum downstream meter tube length, mm
D_p = the difference between the flowing pressure and the equilibrium vapor pressure of the liquid.
D_u = diameter of the meter bore.
ρ = flowing fluid density, kg/m^3
e = orifice edge thickness, mm
E = orifice plate thickness, mm
E_m = modulus of elasticity for steel $[(206.8)(10^6)]$ kPa
F = liquid compressibility factor
F_a = orifice thermal expansion factor. Corrects for the metallic expansion or contraction of the orifice plate. Generally ignored between –20°C and 50°C
F_g = relative density factor applied to change from a relative density of 1.0 (air) to the relative density of the flowing gas
F_{gt} = gravity-temperature factor for liquids
F_c = orifice calculation factor
F_n = numeric conversion factor
F_{na} = units conversion factor for pitot tubes
F_{pb} = pressure base factor applied to change the base pressure from 101.55 kPa (abs)
F_{pm} = pressure factor applied to meter volumes to correct to standard pressure. See API Manual of Petroleum Measurement Standards, Chapter 12, Section 2
F_{pv} = supercompressibility factor required to correct for deviation from the ideal gas laws = $\sqrt{1/Z}$
F_s = steam factor
F_{sl} = orifice slope factor
F_{tb} = temperature base factor. To change the temperature base from 15°C to another desired base
F_{tf} = flowing temperature factor to change from the assumed flowing temperature of 15°C to the actual flowing temperature
F_{tm} = temperature correction factor applied to displacement meter volumes to correct to standard temperature. See API Manual of Petroleum Measurement Standards, Chapter 12, Section 2
G, G_1 = specific gravity at 15°C
G_f = specific gravity at flowing temperature
H = pressure, mm of mercury
h_m = differential pressure measured across the orifice plate in mm of mercury at 15°C
h_w = differential pressure measured across the orifice plate in mm of water at 15°C
$\sqrt{h_w P_f}$ = pressure extension. The square root of the differential pressure times the square root of the absolute static pressure
k = ratio of the specific heat at constant pressure to the specific heat at constant volume
K = a numerical constant. Pulses generated per unit volume through a turbine, positive displacement, coriolis or ultrasonic meter
Key = $F_n (F_c + F_{sl})$ = orifice factor
L = distance between upstream and downstream transducer.
LTB = Length of tube bundle, in flow conditioner, mm (See Fig. 3-3)
MF = meter factor, a number obtained by dividing the actual volume of liquid passed through the meter during proving by the volume registered by the meter
P = pressure, kPa (abs)

FIG. 4-22

Typical Responses Obtained When Determining Ultimate Gain and Ultimate Period

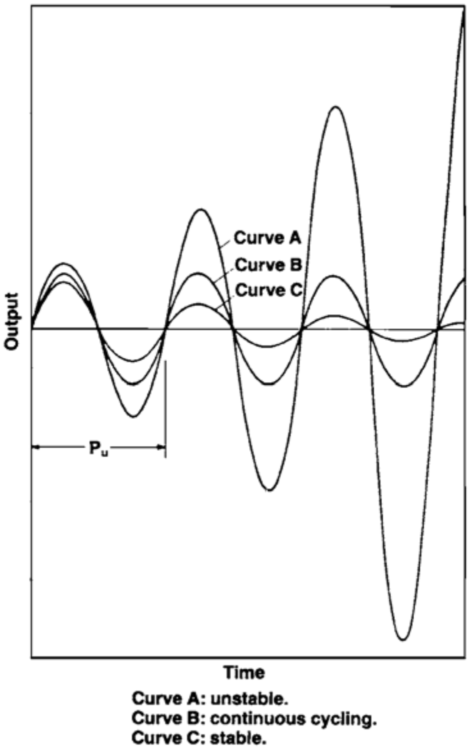


FIG. 4-23

Ziegler-Nichols Settings for 1/4 Decay Response[1]

Mode	K_p	or	PB(%)	T_i	T_d
(P)	0.5 K_u		2(PB_u)	max.	zero
(PI)	0.45 K_u		2.2(PB_u)	$P_u/1.2$	zero
(PD)	0.6 K_u		1.65(PB_u)	max.	$P_u/8.0$
(PID)	0.6 K_u		1.65(PB_u)	$P_u/2.0$	$P_u/8.0$

FIG. 4-24

Typical Controller Settings

Process	Gain	PB(%)	Integral		Derivative
			T_i (sec)	min/repeat	T_d (sec)
Flow	0.6-0.8	167-125	3.0-1.8	0.05-0.03	0.0
Pressure	5.0	20.0	120-60	2.0-1.0	0.0
Temp.	1.0-2.0	100-50	120-30	2.0-0.5	6.0-12
Level	0.8-1.2	125-83	600-300	10.0-5.0	0.6-1.2

Fig. 4-23 gives relative controller gain, integral time, and derivative time for the various control mode combinations for quarter-decay response as related to ultimate controller gain setting, K_u, and ultimate period P_u. Gain settings are also shown in units of proportional band, PB.

Fig. 4-24 shows some typical settings for various types of process controllers.

Example 4-2 — An example using the Ziegler-Nichols method is given below:

For a certain temperature control system, the ultimate sensitivity K_u was found to be 1.5 kPa per °C, and the ultimate period P_u was found to be two minutes. A three mode PID controller is required.

Using Fig. 4-24:

Proportional gain K_p:

$$K_p = 0.6\ K_u = 0.6\ (1.5\ \text{kPa/°C}) = 0.9\ \text{kPa/°C}$$

Integral time constant T_i:

$$T_i = P_u/2,\ T_i = 2/2 = 1.0 \text{ minute}$$

Derivative time constant T_d:

$$T_d = P_u/8,\ T_d = 2/8 = 0.25 \text{ minutes}$$

Control Mode Considerations

The process control engineer has the responsibility for matching the many and variable characteristics of the process to be controlled with the most effective control hardware available.[6] Fig. 4-25 provides guidelines for choosing the mode of control for various types of applications based upon the process reaction rate and size and speed of load changes.

Special considerations should be made in applying a "split-range" controller. A common example is a column temperature controller on a cryogenic demethanizer. In this system the first half (0-50%) of the controller output actuates the "free" heat exchange with the incoming feed, and the second half (50-100%) of the controller output actuates the supplemental heat from the hot oil system. Adaptive gain control may be required since the heating value of the hot oil is much greater than that of the gas used in the heat exchange.

EMBEDDED ADVANCED CONTROL

Embedded advanced control will usually give an improved plant performance over that achievable with traditional techniques. By introducing Embedded Advanced Control, a high level of reliability and security is provided to maximize control system uptime. Since embedded advanced control tools have direct access to controller I/O, they may access process measurements and actuators with no communication jitter or delay. This allows use of these tools on the fastest processes.

CONTROL VALVES

Selecting the proper control valve for each application involves many factors. The valve body design, actuator style, and plug characteristic are critical items for selection. Proper valve sizing is necessary for accurate, efficient, economical process control. In areas where personnel will be affected, noise prediction and control becomes a significant factor.

Engineering application guidelines, nomographs, and equations presented in the following pages may be used to determine the correct control valve configuration, size and flow characteristics, and to predict noise levels for most applications. The material presented here may also be used to evaluate the performance of valves installed in existing plants.

The equations given in this section are used to calculate the flow coefficient (C_v or C_g) required for a valve to pass the re-with the listed Cv should then be used in the chosen sizing equation to calculate a revised, required C_v. This iteration process continues until the calculated C_v and equals the manufactuer's listed C_v.

4. For a new valve selection a valve size is typically chosen such that the maximum, calculated C_v is close to 75% to 85% of valve travel. This allows for process variability while maintaining flow capability. The minimum, calculated C_v should typically occur at or about 10% of valve travel.
5. F_p is the Piping Geometry Factor. It corrects the sizing equations for the effects of fittings such as reducers and expanders that are attached to the valve body ends. F_p values can be determined via test or calculated per the ANSI/ISA S75.01 standard. If the valve has no such fittings attached, e.g., the nominal value size and nominal pipe size are the same, then $F_p = 1.0$. Refer to the full standard for the F_p calculations in cases where fittings do exist.

Other valve configurations, such as ball and butterfly valves, can be sized in a similar manner using the unique X_c and C_v values derived by the manufacturers.

FIG. 4-32

Typical C_v, X_c and F_L Values for Valves*

Valve Style	Body Size, mm	Flow Characteristic					
		Equal Percentage			Linear		
		C_v	X_c	F_L	C_v	X_c	F_L
Globe	25	8	0.74	0.88	17	0.61	0.84
	38	17	0.69	0.84	30	0.70	0.82
	50	25	0.70	0.85	62	0.68	0.77
	63	49	0.66	0.84	84	0.71	0.81
	75	66	0.66	0.82	118	0.70	0.82
	100	125	0.67	0.82	181	0.74	0.82
	150	239	0.74	0.85	367	0.78	0.84
	200	268	0.60	0.85	526	0.74	0.87
Ball	25	16	0.53	0.86	–	–	–
	50	59	0.53	0.81	–	–	–
	75	120	0.50	0.80	–	–	–
	100	195	0.52	0.80	–	–	–
	150	340	0.52	0.80	–	–	–
	200	518	0.54	0.82	–	–	–
	250	1000	0.47	0.80	–	–	–
	300	1530	0.49	0.78	–	–	–
Butterfly	50	60	0.37	0.69	–	–	–
	75	111	0.40	0.69	–	–	–
	100	238	0.40	0.69	–	–	–
	150	635	0.40	0.69	–	–	–
	200	1020	0.40	0.69	–	–	–
	250	1430	0.40	0.69	–	–	–
	300	2220	0.40	0.69	–	–	–
	350	2840	0.40	0.69	–	–	–
	400	3870	0.40	0.69	–	–	–

*At approximately 70% of valve travel. Maximum valve capacity may be estimated using the values given in this figure in conjunction with Fig. 4-29. For a more detailed analysis of capacity capabilities of a given valve at other percentages of travel, consult the valve manufacturer's data.

FIG. 5-20

Fraction of Heat Radiated Values for Flared Gases

Carbon Monoxide	0.075
Hydrogen	0.075
Hydrogen Sulfide	0.070
Ammonia	0.070
Methane	0.10
Propane	0.11
Butane	0.12
Ethylene	0.12
Propylene	0.13

The maximum value of ε for any gas is 0.13.

Spherical Radiation Intensity Formula:

$$I = \frac{(W_f)\,(NHV)\,(\varepsilon)}{14.4\,\pi\,(R^2)} \qquad \textbf{Eq 5-20}$$

This equation has been found to be accurate for distances as close to the flame as one flame length.

Equation 5-20 is valid so long as the proper value of fraction of heat radiated, ε, is inserted. Classically, ε has been considered a fuel property alone. Brzustowski et al.[10] experimentally observed a dependence of ε on jet exit velocity. Other authors have presented models that consider the carbon particle concentration in the flame. The fraction of heat radiated is a function of many variables including gas composition, tip diameter, flare burner design, flowrate and velocity, flame temperature, air-fuel mixing, and steam or air injection; therefore a flare supplier should be consulted to determine the specific values for a given application. A list of vendor recommended fraction of heat radiated values for the most frequently flared gases is shown in Fig. 5.20.

To calculate the intensity of radiation at different locations, it is necessary to determine the length of the flame and its angle in relation to the stack (see Fig. 5-21). A convenient expression to estimate length of flame, L_f, is shown below, based on information from equipment suppliers.

$$L_f = (0.12)\,(d)\sqrt{\frac{\Delta P_w}{1400}} \qquad \textbf{Eq 5-21}$$

or from API 521

$$L_f = 2.14\,(\,Q_r \cdot 10^{-6}\,)^{0.474} \qquad \textbf{Eq 5-22}$$

For conventional (open pipe) flares, an estimate of total flare pressure drop is 1.5 velocity heads based on nominal flare tip diameter. The pressure drop equivalent to 1 velocity head is given by:

$$\Delta P_w = \frac{(0.102)\,\rho\,V^2}{2} = \frac{\rho\,V^2}{19.62} \qquad \textbf{Eq 5-23}$$

ΔP_w is the pressure drop at the tip in mm of water. After determining tip diameter, d, using Eq 5-23, and the maximum required relieving capacity, flame length for conditions other than maximum flow can be calculated using Eq 5-21 and Eq 5-22.

Common practice is to use tip velocities of up to Mach 0.5 for short term emergency flows and Mach 0.2 for maximum continuous flowing.

$$d = \left(\sqrt{\frac{3.23 \cdot 10^{-5} \cdot W}{P_2 \bullet M} \cdot \left(\frac{Z \cdot T}{k \cdot MW}\right)^{0.5}}\right) \cdot 1000 \qquad \textbf{Eq 5-24}$$

Sonic velocity of a gas is given by:

$$a = \sqrt{k\,\frac{R_o}{MW}\,T} \qquad \textbf{Eq 5-25}$$

The center of the flame is assumed to be located at a distance equal to 1/3 the length of the flame from the tip.

The angle of the flame results from the vectorial addition of the velocity of the wind and the gas exit velocity.

$$\theta = \tan^{-1}\left(\frac{V_w}{V_{ex}}\right) \qquad \textbf{Eq 5-26}$$

$$V_{ex} = 168\sqrt{\frac{\Delta P_W}{1400}} \qquad \textbf{Eq 5-27}$$

Note: API gives a greater lean angle

The coordinates of the flame center with respect to the tip are:

$$X_c = (L_f\,/\,3)\,(\sin\theta) \qquad \textbf{Eq 5-28}$$

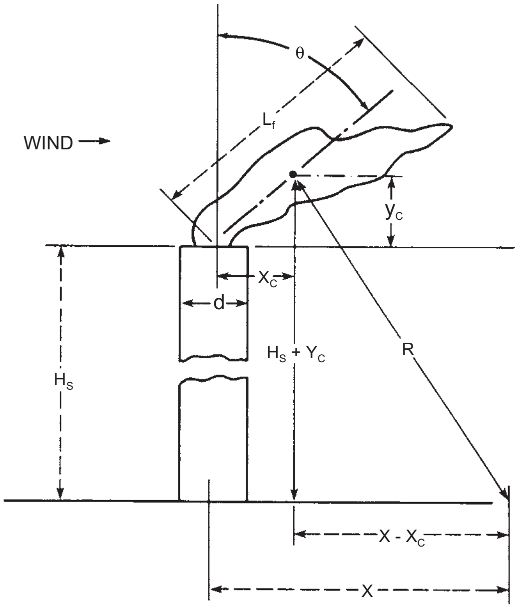
$$Y_c = (L_f\,/\,3)\,(\cos\theta) \qquad \textbf{Eq 5-29}$$

The distance from any point on the ground level to the center of the flame is:

$$R = \sqrt{(X - X_c)^2 + (H_s + Y_c)^2} \qquad \textbf{Eq 5-30}$$

FIG. 5-21

Dimensional References for Sizing a Flare Stack



Courtesy American Petroleum Institute

FIG. 6-26
Partial Volumes of Spheres — Cubic Meters

Diam. of Sphere	Depth of Liquid, meters																	
meters	0.5	1	2	4	6	8	10	12	14	16	18	20	25	30	35	40	45	50
0.5	0.065	–																
1	0.262	0.524	–															
2	0.654	2.094	4.189	–														
4	1.440	5.236	16.755	33.510	–													
6	2.225	8.378	29.322	83.776	113.097	–												
8	3.011	11.519	41.888	134.041	226.194	268.082	–											
10	3.796	14.661	54.454	184.307	339.292	469.144	523.598	–										
12	4.581	17.802	67.021	234.572	452.389	670.206	837.757	904.778	–									
14	5.367	20.944	79.587	284.837	565.486	871.268	1 151.916	1 357.167	1 436.754	–								
16	6.152	24.086	92.153	335.103	678.583	1 072.329	1 466.075	1 809.556	2 052.505	2 144.66	–							
18	6.938	27.227	104.720	385.368	791.681	1 273.391	1 780.234	2 261.945	2 668.257	2 948.91	3 053.63	–						
20	7.723	30.369	117.286	435.634	904.778	1 474.453	2 094.393	2 714.334	3 284.009	3 753.15	4 071.50	4 188.79	–					
25	9.687	38.223	148.702	561.297	1 187.521	1 977.107	2 879.791	3 845.306	4 823.388	5 763.77	6 616.19	7 330.38	8 181.22	–				
30	11.650	46.077	180.118	686.961	1 470.264	2 479.762	3 665.188	4 976.279	6 362.767	7 774.39	9 160.88	10 471.97	13 089.96	14 137.16	–			
35	13.614	53.931	211.534	812.625	1 753.007	2 982.416	4 450.586	6 107.251	7 902.146	9 785.01	11 705.56	13 613.56	17 988.69	21 205.73	22 449.28	–		
40	15.577	61.785	242.950	938.288	2 035.750	3 485.071	5 235.983	7 238.223	9 441.525	11 795.62	14 250.25	16 755.15	22 907.43	28 274.31	32 070.40	33 510.29	–	
45	17.541	69.639	274.366	1 063.952	2 318.493	3 987.725	6 021.381	8 369.196	10 980.904	13 806.24	16 794.94	19 896.74	27 816.16	35 342.89	41 691.52	46 076.65	47 712.90	–
50	19.504	77.493	305.781	1 189.615	2 601.237	4 490.379	6 806.778	9 500.168	12 520.283	15 816.86	19 339.63	23 038.33	32 724.90	42 411.47	51 312.64	58 643.01	63 617.20	65 449.79

FIG. 6-27
Approximate Contents (Cubic Meters) of Rectangular Tanks Per Meter of Liquid*

Tank Width, m	Tank Length, m							
	1.0	2.0	3.0	4.0	5.0	6.0	7.0	8.0
0.5	0.50	1.00	1.50	2.00	2.50	3.00	3.50	4.00
1.0	1.00	2.00	3.00	4.00	5.00	6.00	7.00	8.00
1.5	1.50	3.00	4.50	6.00	7.50	9.00	10.50	12.00
2.0	2.00	4.00	6.00	8.00	10.00	12.00	14.00	16.00
2.5	2.50	5.00	7.50	10.00	12.50	15.00	17.50	20.00
3.0	3.00	6.00	9.00	12.00	15.00	18.00	21.00	24.00
3.5	3.50	7.00	10.50	14.00	17.50	21.00	24.50	28.00
4.0	4.00	8.00	12.00	16.00	20.00	24.00	28.00	32.00
4.5	4.50	9.00	13.50	18.00	22.50	27.00	31.50	36.00
5.0	5.00	10.00	15.00	20.00	25.00	30.00	35.00	40.00
5.5	5.50	11.00	16.50	22.00	27.50	33.00	38.50	44.00
6.0	6.00	12.00	18.00	24.00	30.00	36.00	42.00	48.00
6.5	6.50	13.00	19.50	26.00	32.50	39.00	45.50	52.00
7.0	7.00	14.00	21.00	28.00	35.00	42.00	49.00	56.00
7.5	7.50	15.00	22.50	30.00	37.50	45.00	52.50	60.00
8.0	8.00	16.00	24.00	32.00	40.00	48.00	56.00	64.00
8.5	8.50	17.00	25.50	34.00	42.50	51.00	59.50	68.00
9.0	9.00	18.00	27.00	36.00	45.00	54.00	63.00	72.00
9.5	9.50	19.00	28.50	38.00	47.50	57.00	66.50	76.00
10.0	10.00	20.00	30.00	40.00	50.00	60.00	70.00	80.00
10.5	10.50	21.00	31.50	42.00	52.50	63.00	73.50	84.00
11.0	11.00	22.00	33.00	44.00	55.00	66.00	77.00	88.00
11.5	11.50	23.00	34.50	46.00	57.50	69.00	80.50	92.00
12.0	12.00	24.00	36.00	48.00	60.00	72.00	84.00	96.00

1 cu meter = 264.172 U.S. gal.
= 219.9692 Imperial gallons
= 6.2898 bbls (42 U.S. gals)

SECTION 7

Separation Equipment

PRINCIPLES OF SEPARATION

Three principles used to achieve physical separation of gas and liquids or solids are momentum, gravity settling, and coalescing. Any separator may employ one or more of these principles, but the fluid phases must be "immiscible" and have different densities for separation to occur.

FIG. 7-1

Nomenclature

A = area, m^2
A_p = particle or droplet cross sectional area, m^2
C = empirical constant for separator sizing, m/h
C^* = empirical constant for liquid-liquid separators, $(m^3 \bullet mPa \bullet s)/(m^2 \bullet day)$
C' = drag coefficient of particle, dimensionless (Fig. 7-3)
D_i = separator inlet nozzle diameter, mm
D_p = droplet diameter, m
D_v = inside diameter of vessel, mm
G_m = maximum allowable gas mass-velocity necessary for particles of size D_p to drop or settle out of gas, $kg/(h \bullet m^2)$
g = acceleration due to gravity, 9.81 m/s^2
H_l = width of liquid interface area, m
J = gas momentum, $kg/(m \bullet s^2)$
K = empirical constant for separator sizing, m/s
K_{CR} = proportionality constant from Fig. 7-5 for use in Eq 7-5, dimensionless
L = seam to seam length of vessel, mm
L_l = length of liquid interface, mm
M = mass flow, kg/s
M_p = mass of droplet or particle, kg
MW = molecular mass, kg/(kg mole)
P = system pressure, kPa(abs)
Q = estimated gas flow capacity, $(Sm^3/day)/m^2$ of filter area
Q_A = actual gas flow rate, m^3/s
R = gas constant, 8.31 $[kPa(abs) \bullet m^3]/[K \bullet kg\ mole]$
Re = Reynolds number, dimensionless
S_{hl} = relative density of heavy liquid, water = 1.0
S_{ll} = relative density of light liquid, water = 1.0
T = system temperature, K
t = retention time, minutes
U = volume of settling section, m^3
V_t = critical or terminal gas velocity necessary for particles of size D_p to drop or settle out of gas, m/s
W = total liquid flow rate, m^3/day
W_{cl} = flow rate of light condensate liquid, m^3/day
Z = compressibility factor, dimensionless

Greek:

ρ_g = gas phase density, kg/m^3
ρ_l = liquid phase density, droplet or particle, kg/m^3
μ = viscosity of continuous phase, $mPa \bullet s$

Filter Separators: A filter separator usually has two compartments. The first compartment contains filter-coalescing elements. As the gas flows through the elements, the liquid particles coalesce into larger droplets and when the droplets reach sufficient size, the gas flow causes them to flow out of the filter elements into the center core. The particles are then carried into the second compartment of the vessel (containing a vane-type or knitted wire mesh mist extractor) where the larger droplets are removed. A lower barrel or boot may be used for surge or storage of the removed liquid.

Flash Tank: A vessel used to separate the gas evolved from liquid flashed from a higher pressure to a lower pressure.

Line Drip: Typically used in pipelines with very high gas-to-liquid ratios to remove only free liquid from a gas stream, and not necessarily all the liquid. Line drips provide a place for free liquids to separate and accumulate.

Liquid-Liquid Separators: Two immiscible liquid phases can be separated using the same principles as for gas and liquid separators. Liquid-liquid separators are fundamentally the same as gas-liquid separators except that they must be designed for much lower velocities. Because the difference in density between two liquids is less than between gas and liquid, separation is more difficult.

Scrubber or Knockout: A vessel designed to handle streams with high gas-to-liquid ratios. The liquid is generally entrained as mist in the gas or is free-flowing along the pipe wall. These vessels usually have a small liquid collection section. The terms are often used interchangeably.

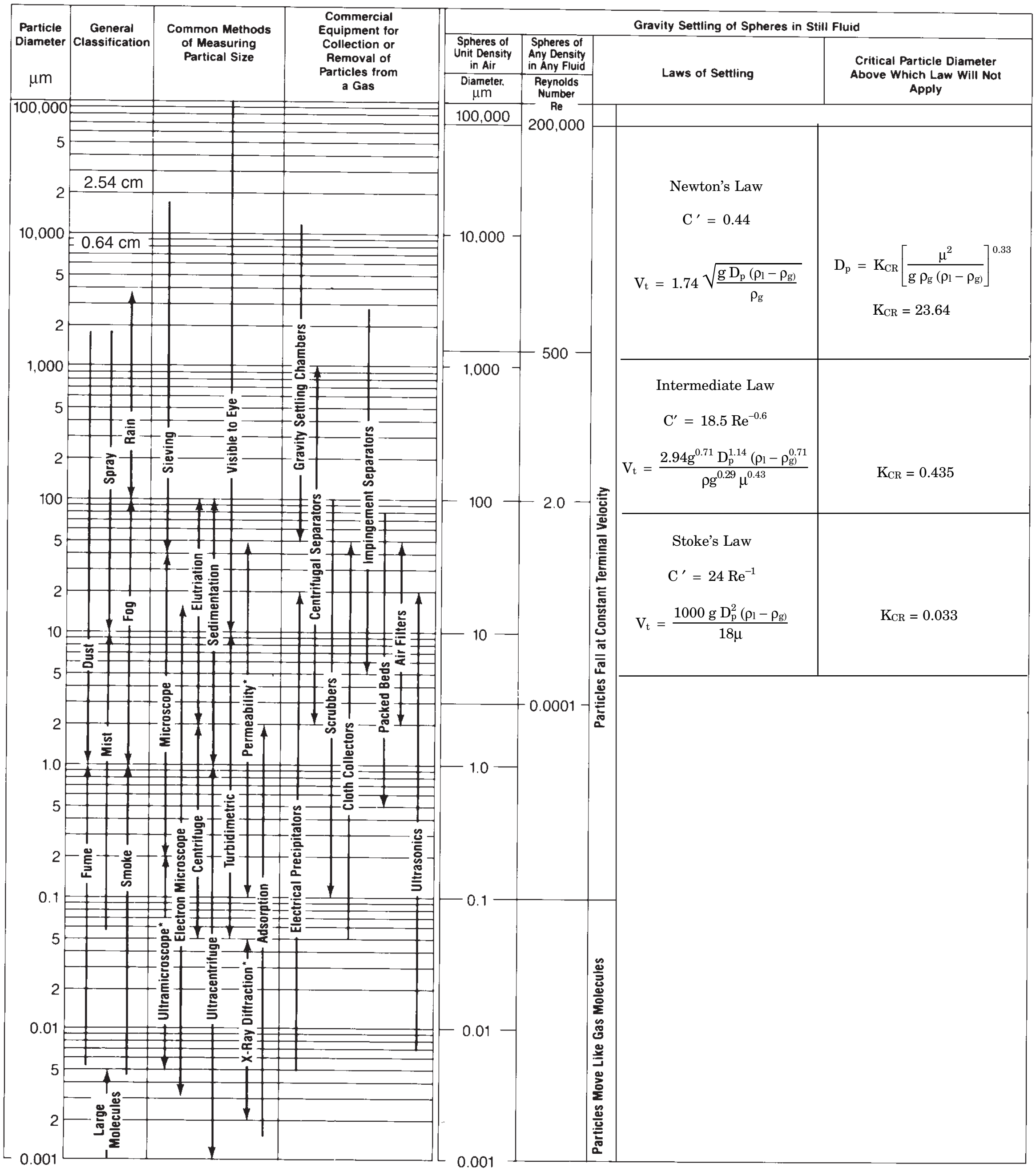
Separator: A vessel used to separate a mixed-phase stream into gas and liquid phases that are "relatively" free of each other. Other terms used are scrubbers, knockouts, line-drips, and decanters.

Slug Catcher: A particular separator design able to absorb sustained in-flow of large liquid volumes at irregular intervals. Usually found on gas gathering systems or other two-phase pipeline systems. A slug catcher may be a single large vessel or a manifolded system of pipes.

Three Phase Separator: A vessel used to separate gas and two immiscible liquids of different densities (e.g. gas, water, and oil).

FIG. 7-5

Gravity Settling Laws and Particle Characteristics



Adapted From Ref. 7-3

Note that except for the Natural gas – condensate application, the residence times specified in Fig. 7-20 are primarily based on process control stability/operability.

API 12J gives the following guidelines for gas-oil separation

These values are primarily intended to reflect liquid degassing requirements. In practice, process control stability and operability requirements will often override the degassing requirements.

Vessel layout recommendations, including liquid handling requirements, are given in Fig. 7-8 and 7-9 for vertical and horizontal separators, respectively.

The retention time requirements given in Figures 7-20 and 7-21 are not specific to vessel orientation. However, the liquid degassing process actually involves the separation of gas bubbles from the liquid phase, which under ideal conditions can be described by the gravity settling equation, Eq 7-1. Similar to liquid droplet settling out of the gas phase, it is easier for a gas bubble to rise perpendicularly through the moving liquid in a horizontal separator than directly against the downflowing liquid in a vertical vessel. Theoretically, for equal liquid residence times, the horizontal separator should be slightly more efficient at degassing. However, this has not typically been an issue in practice. If it is deemed necessary to calculate vessel liquid handling requirements for a degassing constraint according to gravity settling theory, a gas bubble size of 150–200 μm has been suggested by several sources.

FIG. 7-20

Typical Retention Times for Gas/Liquid Separator

Application	Retention Time, minutes
Natural Gas – Condensate separation	2 – 4
Fractionator Feed Tank	10 – 15
Reflux Accumulator	5 – 10
Fractionation Column Sump	2
Amine Flash Tank	5 – 10
Refrigeration Surge Tank	5
Refrigeration Economizer	3
Heat Medium Oil Surge Tank	5 – 10

FIG. 7-21

API 12J Retention Times for Oil-Gas Separators

Oil Gravity	Liquid Retention Time, min
> 35	1
20 – 35	1 to 2
10 – 20	2 to 4

Three-Phase and Liquid-Liquid Separation

The gas handling requirements for three-phase separation are dealt with in a similar manner as discussed for two-phase separation. Traditionally, sizing for liquid-liquid separation has involved specification of liquid residence times.

Fig. 7-22 provides suggested residence times for various liquid-liquid separation applications. These figures generally assume equal residence times for both the light and heavy liquid phases.

While the residence time approach for liquid-liquid separation equipment design has been widely used in industry for years, it does have some limitations.

- the typical approach of assuming equal residence times for both liquid phases may not be optimum, e.g. It is generally much easier to separate oil droplets from water than vice-versa. Settling theory (Eq 7-1) explains this as being due to the lower viscosity of water compared to oil.
- Residence times do not take into account vessel geometry, i.e. 3 minutes residence time in the bottom of a tall, small diameter vertical vessel will not achieve the same separation performance as 3 minutes in a horizontal separator, again according to droplet settling theory.
- The residence time method does not provide any direct indication as to the quality of the separated liquids, e.g. amount of water in the hydrocarbon or the amount of hydrocarbon in the water. Droplet settling theory can not do this either in most cases, but there is some empirical data available which allows for approximate predictions in specific applications.

Removal of very small droplets may require the use of specialized internals or the application of electrostatic fields to promote coalescence.

Liquid-liquid separation may be divided into two broad categories of operation. The first is defined as "gravity separation,"

FIG. 7-22

Typical Retention Times for Liquid-Liquid Separation

Type of Separation	Retention Time, minutes
Hydrocarbon/Water Separators[7]	
Above 35° API hydrocarbon	3 – 5
Below 35° API hydrocarbon	
100°F and above	5 – 10
80°F	10 – 20
60°F	20 – 30
Ethylene Glycol/Hydrocarbon[8] Separators (Cold Separators)	20 – 60
Amine/Hydrocarbons Separators[9]	20 – 30
Coalescer, Hydrocarbon/Water Separators12	
100°F and above	5 – 10
80°F	10 – 20
60°F	20 – 30
Caustic/Propane	30 – 45
Caustic/Heavy Gasoline	30 – 90

SECTION 12

Pumps & Hydraulic Turbines

Pumps

The most common types of pumps used in gas processing plants are centrifugal and positive displacement. Occasionally regenerative turbine pumps, axial-flow pumps, and ejectors are used.

Modern practice is to use centrifugal rather than positive displacement pumps where possible because they are usually less costly, require less maintenance, and less space. Conventional centrifugal pumps operate at speeds between 1200 and 8000 rpm. Very high speed centrifugal pumps, which can operate

FIG. 12-1

Nomenclature

A = cross-sectional area of plunger, piston, or pipe, mm^2
a = cross-sectional area of piston rod, mm^2
AC = alternating current
bbl = barrel (42 U.S. gallons or 0.1589 m^3)
bkW = brake kilowatt
C = constant (Fig. 12-19)
C_p = specific heat at average temperature, J/(kg • °C)
D = displacement of reciprocating pump, m^3/h
DC = direct current
d = impeller diameter, mm
e = pump efficiency, fraction
g = 9.8067 m/s^2 (acceleration of gravity)
H = total equipment head, m of fluid
h = head, m of fluid pumped
hyd kW = hydraulic kilowatts
k = factor related to fluid compressibility (Fig. 12-19)
K = type of pump factor (Eq 12-18)
kPa = kilopascal
kPa (abs) = kilopascal, absolute
kPa (ga) = kilopascal, gage
L = length of suction pipe, m
L_s = stroke length, mm
m = number of plungers or pistons
NPSH = net positive suction head of fluid pumped, m
NPSHA = NPSH available, m
NPSHR = NPSH required, m
n = speed of rotation, revolutions/minute (rpm)
n_s = specific speed (See Fig. 12-2 for units)
N = Polytropic exponent of charge gas. (For nitrogen, N = 1.4)
ΔP = differential pressure, kPa
P = pressure, kPa (abs) or kPa (ga)
P_{vp} = liquid vapor pressure at pumping temperature, kPa (abs)
Q = rate of liquid flow, m^3/h
r = ratio of internal volume of fluid between valves, when the piston or plunger is at the end of the suction stroke, to the piston or plunger displacement.
RD = relative density to water at standard temperature
s = slip or leakage factor for reciprocating and rotary pumps
S = suction specific speed (units per Eq 12-7)
sp gr = specific gravity at average flowing conditions. Equal to RD
T = torque, N • m (Newton meters)
t_r = temperature rise, °C
u = impeller peripheral velocity, m/s
VE = volumetric efficiency, fraction
VE_o = overall volumetric efficiency
VE_ρ = volumetric efficiency due to density change
VE_l = volumetric efficiency due to leakage
V_{pd} = pulsation dampener volume, mm^3
v = liquid mean velocity at a system point, m/s
z = elevation of a point of the system above (+) or below (–) datum of the pump. For piping, the elevation is from the datum to the piping centerline; for vessels and tanks, the elevation is from the datum to the liquid level.

Greek:

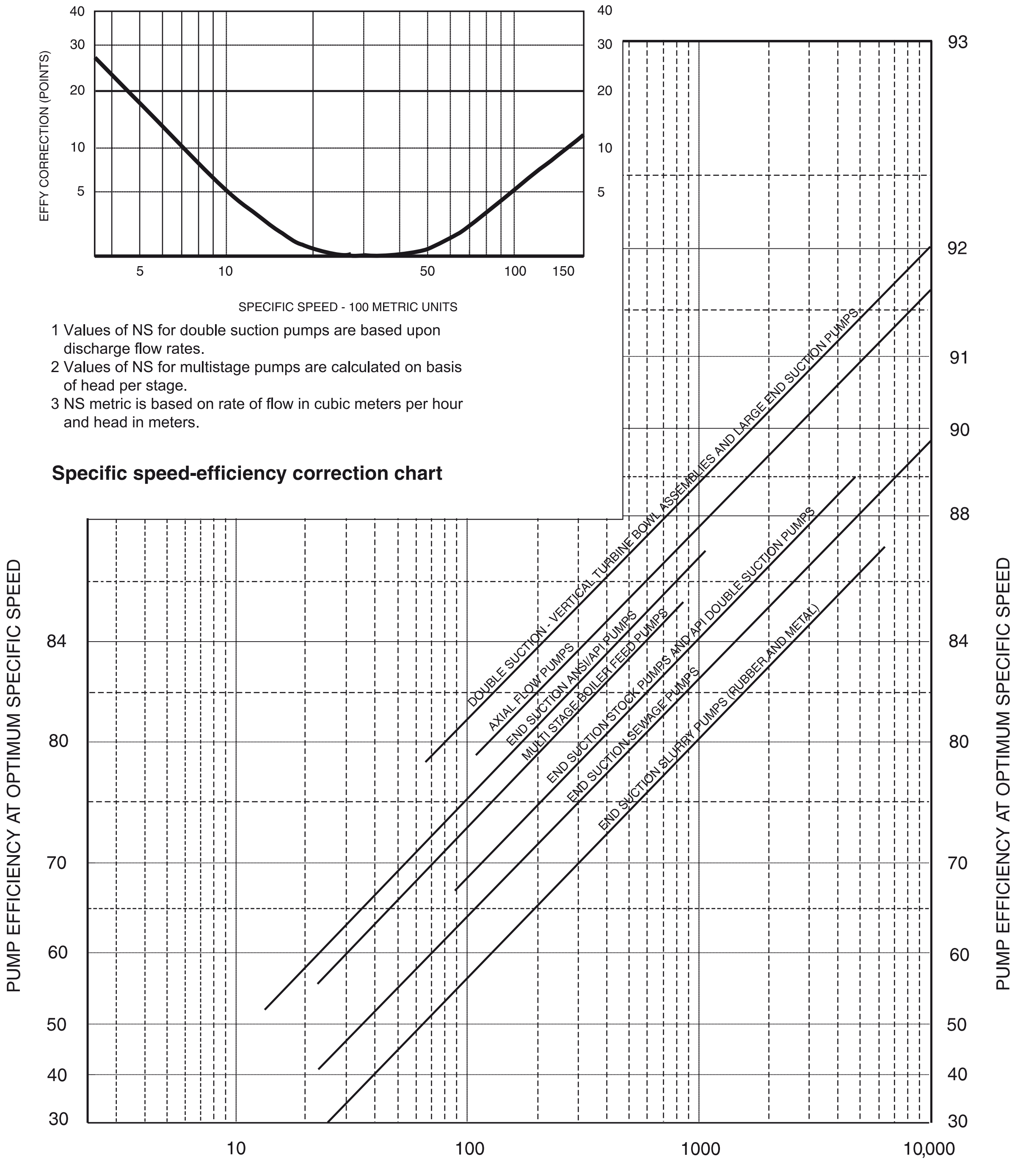
ρ = density at average flowing conditions, kg/m^3
ρ_i = inlet density, kg/m^3
ρ_o = outlet density, kg/m^3
Δ = allowable pressure fluctuations as a percentage of mean pressure

Subscripts:

a = acceleration
ave - with P, average pressure in pulsating flow
bep = best efficiency point, for maximum impeller diameter
c = compression
d = discharge of pump
dv = discharge vessel
D = displacement
f = friction
i = inlet of equipment
l = leakage
max = with P, maximum acceptable peak pressure in pulsating flow
min = with P, minimum acceptable valley pressure in pulsating flow
o = outlet of equipment
ov = overall
p = pressure
r = rise

Fig. 12-13

Optimum Generally Attainable Efficiency Chart



1 Values of NS for double suction pumps are based upon discharge flow rates.
2 Values of NS for multistage pumps are calculated on basis of head per stage.
3 NS metric is based on rate of flow in cubic meters per hour and head in meters.

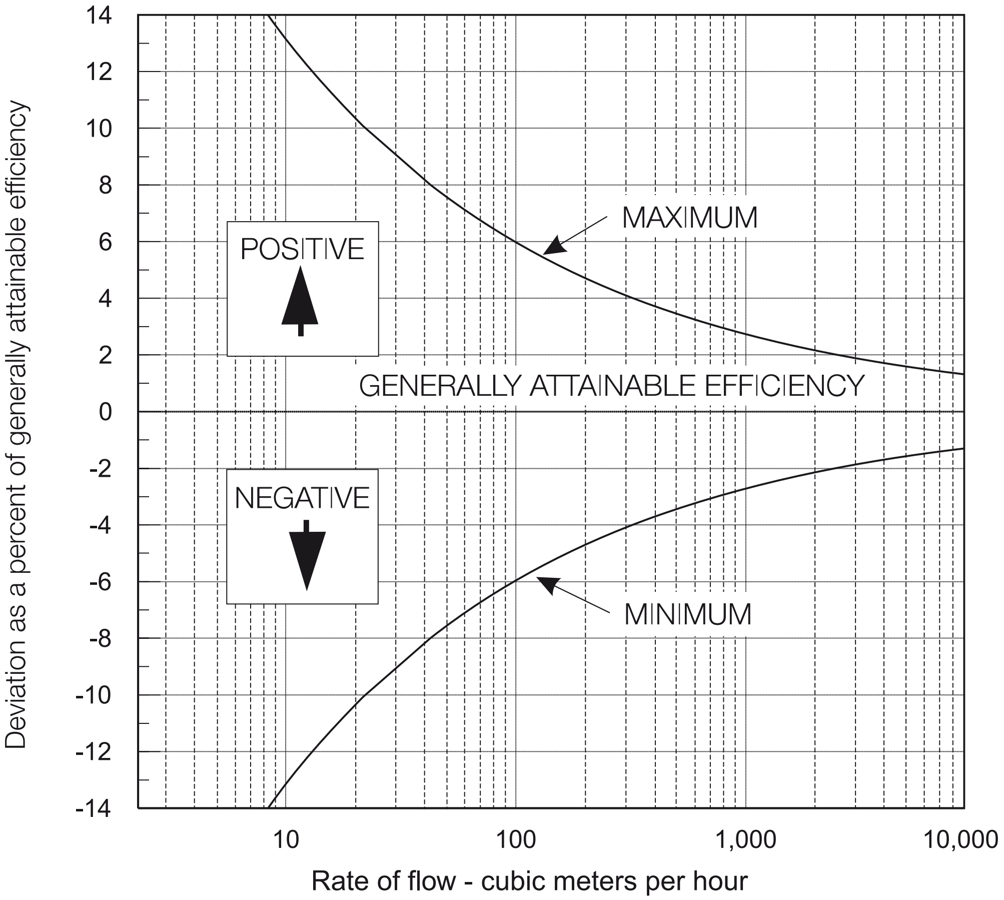
Specific speed-efficiency correction chart

NOTES:
1 The above charts depict the generally attainable efficiency levels of centrifugal pumps at best efficiency point with maximum diameter impeller when pumping clear water at 30°C (85°F).
2 The above charts relate to industrial class pumps designed, manufactured and tested in accordance with recognized industry standards.

Courtesy of the Hydraulic Institute, www.pumps.org

Fig. 12-14

Deviation from Generally Attainable Efficiency



Courtesy of the Hydraulic Institute, www.pumps.org

unnecessary oversizing of the motor, an alternate proposal shall be submitted for the purchaser's approval."

Motor Nameplate Rating		Percentage of Rated Pump Power
kW	hp	
<22	<30	125
22-55	30-75	115
>55	>75	110

Alternatives to electric motor drivers are:

- internal combustion engines
- gas turbines
- steam turbines
- hydraulic power-recovery turbines

Usually the speed of rotation of these drivers can be varied to provide control.

Variable Speed Drives — Fig. 12-18 lists various types of adjustable speed drives, their characteristics and their application.

Materials of Construction

Pumps manufactured with cast-steel cases and cast-iron internals are most common in the gas processing industry. API Std 610 is a good reference for material selection. The material selections in this document can be over-ridden as required to reflect experience.

Experience is the best guide to selection of materials for pumps. Process pump manufacturers can usually provide suggestions for materials, based on their experience and knowledge of pumps.

Shaft Seals

Mechanical seals are the most common sealing devices for centrifugal pumps in process service. The purpose of the seal is to retain the pumped liquid inside the pump at the point

FIG. 12-17

Check List for Centrifugal Pump Troubles and Causes

Trouble:	Possible Causes:
1. Failure to deliver liquid	a. Wrong direction of rotation b. Pump not primed c. Suction line not filled with liquid d. Air or vapor pocket in suction line e. Inlet to suction pipe not sufficiently submerged f. Available NPSH not sufficient g. Pump not up to rated speed h. Total head required greater than head which pump is capable of delivering
2. Pump does not deliver rated capacity	a. Wrong direction of rotation b. Suction line not filled with liquid c. Air or vapor pocket in suction line d. Air leaks in suction line or stuffing boxes e. Inlet to suction pipe not sufficiently submerged. f. Available NPSH not sufficient g. Pump not up to rated speed h. Total head greater than head for which pump designed j. Foot valve too small k. Foot valve clogged with trash m. Viscosity of liquid greater than that for which pump designed n. Mechanical defects: (1) Wearing rings worn (2) Impeller damaged (3) Internal leakage resulting from defective gaskets o. Discharge valve not fully opened
3. Pump does not develop rated discharge pressure	a. Gas or vapor in liquid b. Pump not up to rated speed c. Discharge pressure greater than pressure for which pump designed d. Viscosity of liquid greater than that for which pump designed e. Wrong rotation f. Mechanical defects: (1) Wearing rings worn (2) Impeller damaged (3) Internal leakage resulting from defective gaskets
4. Pump loses liquid after starting	a. Suction line not filled with liquid b. Air leaks in suction line or stuffing boxes c. Gas or vapor in liquid d. Air or vapor pockets in suction line e. Inlet to suction pipe not sufficiently submerged f. Available NPSH not sufficient g. Liquid seal piping to lantern ring plugged h. Lantern ring not properly located in stuffing box
5. Pump overloads driver	a. Speed too high b. Total head lower than rated head c. Excessive recirculation d. Either or both the relative density and viscosity of liquid different from that for which pump is rated e. Mechanical defects: (1) Misalignment (2) Shaft bent (3) Rotating element dragging (4) Packing too tight
6. Vibration	a. Starved suction (1) Gas or vapor in liquid (2) Available NPSH not sufficient (3) Inlet to suction line not sufficiently submerged (4) Gas or vapor pockets in suction line b. Misalignment c. Worn or loose bearings d. Rotor out of balance (1) Impeller plugged (2) Impeller damaged e. Shaft bent f. Improper location of control valve in discharge line g. Foundation not rigid
7. Stuffing boxes overheat	a. Packing too tight b. Packing not lubricated c. Wrong grade of packing d. Insufficient cooling water to jackets e. Box improperly packed.
8. Bearings overheat	a. Oil level too low b. Improper or poor grade of oil c. Dirt in bearings d. Dirt in oil e. Moisture in oil f. Oil cooler clogged or scaled g. Failure of oiling system h. Insufficient cooling water circulation i. Insufficient cooling air k. Bearings too tight m. Oil seals too close fit on shaft n. Misalignment
9. Bearings wear rapidly	a. Misalignment b. Shaft bent c. Vibration d. Excessive thrust resulting from mechanical failure inside the pump e. Lack of lubrication f. Bearings improperly installed g. Dirt in bearings h. Moisture in oil j. Excessive or insufficient cooling of bearings

For double acting cylinders, the percent clearance is based on the total clearance volume for both the head end and the crank end of a cylinder. These two clearance volumes are not the same due to the presence of the piston rod in the crank end of the cylinder. Sometimes additional clearance volume (external) is intentionally added to reduce cylinder capacity.

The term "volumetric efficiency" refers to the actual pumping capacity of a cylinder compared to the piston displacement. Without a clearance volume for the gas to expand and delay the opening of the suction valve(s), the cylinder could deliver its entire piston displacement as gas capacity. The effect of the gas contained in the clearance volume on the pumping capacity of a cylinder can be represented by:

$$VE = 100 - r - C\left[\frac{Z_s}{Z_d}(r^{1/k}) - 1\right] \qquad \textbf{Eq 13-14}$$

Volumetric efficiencies as determined by Eq. 13-14 are theoretical in that they do not account for suction and discharge valve losses. The suction and discharge valves are actually spring-loaded check valves that permit flow in one direction only. The springs require a small differential pressure to open. For this reason, the pressure within the cylinder at the end of the suction stroke is lower than the line suction pressure and, likewise, the pressure at the end of the discharge stroke is higher than line discharge pressure.

One method for accounting for suction and discharge valve losses is to reduce the volumetric efficiency by an arbitrary amount, typically 4%, thus modifying Eq. 13-14 as follows:

$$VE = 96 - r - C\left[\frac{Z_s}{Z_d}(r^{1/k}) - 1\right] \qquad \textbf{Eq 13-15}$$

When a non-lubricated compressor is used, the volumetric efficiency should be corrected by subtracting an additional 5% for slippage of gas. This is a capacity correction only and, as a first approximation, would not be considered when calculating compressor horsepower. The energy of compression is used by the gas even though the gas slips by the rings and is not discharged from the cylinder.

If the compressor is in propane, or similar heavy gas service, an additional 4% should be subtracted from the volumetric efficiency. These deductions for non-lubricated and propane performance are both approximate and, if both apply, cumulative.

Fig. 13-10 provides the solution to the function $r^{1/k}$. Values for compression ratios not shown may be obtained by interpo-

FIG. 13-10

Values of $r^{1/k}$

Compression Ratio	k, isentropic exponent C_p/C_v								
	1.10	**1.14**	**1.18**	**1.22**	**1.26**	**1.30**	**1.34**	**1.38**	**1.42**
1.2	1.180	1.173	1.167	1.161	1.156	1.151	1.146	1.141	1.137
1.4	1.358	1.343	1.330	1.318	1.306	1.295	1.285	1.276	1.267
1.6	1.533	1.510	1.489	1.470	1.452	1.436	1.420	1.406	1.392
1.8	1.706	1.675	1.646	1.619	1.594	1.572	1.551	1.531	1.513
2.0	1.878	1.837	1.799	1.765	1.733	1.704	1.677	1.652	1.629
2.2	2.048	1.997	1.951	1.908	1.870	1.834	1.801	1.771	1.742
2.4	2.216	2.155	2.100	2.050	2.003	1.961	1.922	1.886	1.852
2.6	2.384	2.312	2.247	2.188	2.135	2.086	2.040	1.999	1.960
2.8	2.550	2.467	2.393	2.326	2.264	2.208	2.156	2.109	2.065
3.0	2.715	2.621	2.537	2.461	2.391	2.328	2.270	2.217	2.168
3.2	2.879	2.774	2.680	2.595	2.517	2.447	2.382	2.323	2.269
3.4	3.042	2.926	2.821	2.727	2.641	2.563	2.492	2.427	2.367
3.6	3.204	3.076	2.961	2.857	2.764	2.679	2.601	2.530	2.465
3.8	3.366	3.225	3.100	2.987	2.885	2.792	2.708	2.631	2.560
4.0	3.526	3.374	3.238	3.115	3.005	2.905	2.814	2.731	2.655
4.2	3.686	3.521	3.374	3.242	3.124	3.016	2.918	2.829	2.747
4.4	3.846	3.668	3.510	3.368	3.241	3.126	3.021	2.926	2.839
4.6	4.004	3.814	3.645	3.493	3.357	3.235	3.123	3.022	2.929
4.8	4.162	3.959	3.779	3.617	3.473	3.342	3.224	3.116	3.018
5.0	4.319	4.103	3.912	3.740	3.587	3.449	3.324	3.210	3.106
5.2	4.476	4.247	4.044	3.863	3.700	3.554	3.422	3.303	3.193
5.4	4.632	4.390	4.175	3.984	3.813	3.659	3.520	3.394	3.279
5.6	4.788	4.532	4.306	4.105	3.925	3.763	3.617	3.485	3.364
5.8	4.943	4.674	4.436	4.224	4.035	3.866	3.713	3.574	3.448
6.0	5.098	4.815	4.565	4.343	4.146	3.968	3.808	3.663	3.532
6.2	5.252	4.955	4.694	4.462	4.255	4.069	3.902	3.751	3.614
6.4	5.406	5.095	4.822	4.579	4.363	4.170	3.996	3.839	3.696
6.6	5.560	5.235	4.949	4.696	4.471	4.270	4.089	3.925	3.777
6.8	5.713	5.374	5.076	4.813	4.578	4.369	4.181	4.011	3.857
7.0	5.865	5.512	5.202	4.928	4.685	4.468	4.272	4.096	3.937

Gas Pulsation Control

Pulsation is inherent in reciprocating compressors because suction and discharge valves are open during only part of the stroke.

Pulsation must be damped (controlled) in order to:

a. provide smooth flow of gas to and from the compressor,
b. prevent overloading or underloading of the compressors, and
c. reduce overall vibration.

There are several types of pulsation chambers. The simplest one is a volume bottle, or a surge drum, which is a pressure vessel, unbaffled internally and mounted on or very near a cylinder inlet or outlet.

A manifold joining the inlet and discharge connections of cylinders operating in parallel can also serve as a volume bottle.

Performance of volume bottles is not normally guaranteed without an analysis of the piping system from the compressor to the first process vessel.

Volume bottles are sized empirically to provide an adequate volume to absorb most of the pulsation. Several industry methods were tried in an effort to produce a reasonable rule-of-thumb for their sizing. Fig. 13-20 may be used for approximate bottle sizing.

Example 13-3

Indicated suction pressure = 4000 kPa (abs)

Indicated discharge pressure = 9500 kPa (abs)

Cylinder bore = 150 mm

Cylinder stroke = 380 mm

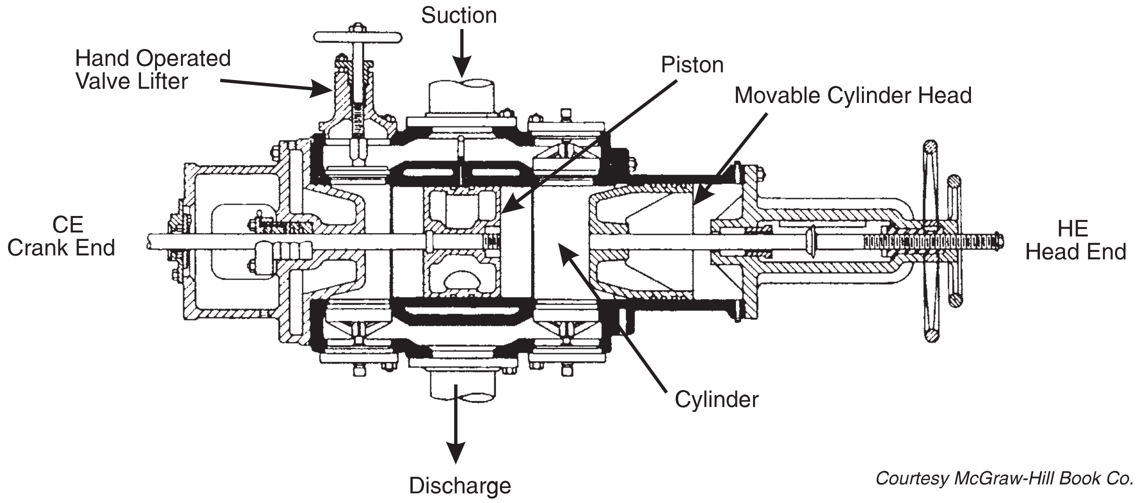
Swept volume = π ($150^2/4$) (380) = 6 715 154 mm^3 = 0.006715 m^3

From Fig. 13-20:

At 4000 kPa inlet pressure, the suction bottle multiplier is approximately 7.5. Suction-bottle volume = (7.5) (6 715 154) = 50 363 657 mm^3 = 0.0504 m^3

FIG. 13-19

Sectional View of a Cylinder Equipped with a Hand-Operated Valve Lifter and Variable-Volume Clearance



Courtesy McGraw-Hill Book Co.

FIG. 13-20

Approximate Bottle Sizing Chart

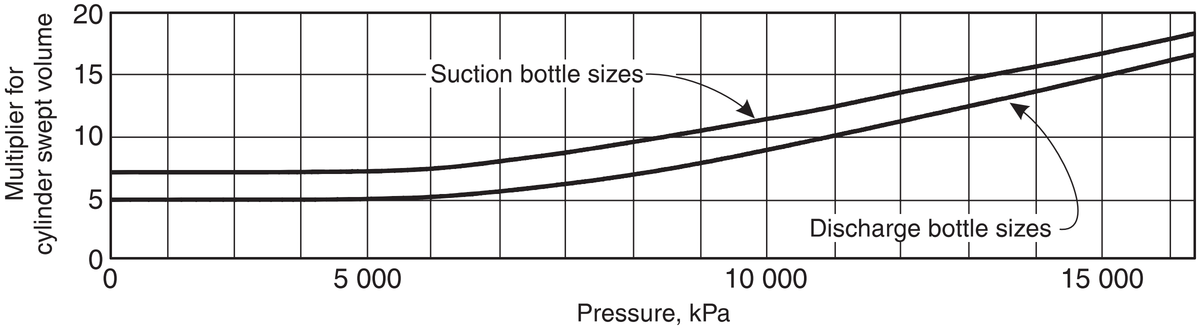


FIG. 18-14

Types of Ion-Exchange Processes

(A)	$Ca(HCO_3)_2$	→	Cation	→	$NaHCO_3$
			Na^{+1}		
	$CaSO_4$	→	Exchanger	→	Na_2SO_4
(B)	$Ca(HCO_3)_2$	→	Cation	→	H_2CO_3
			H^{+1}		
	$CaSO_4$	→	Exchanger	→	H_2SO_4
(C)	Na_2SO_4	→	Anion	→	NaCl
			Cl^{-1}		
	$NaHCO_3$	→	Exchanger	→	NaCl
(D)	H_2CO_3	→	Anion	→	H_2O
			OH^{-1}		
	H_2SO_4	→	Exchanger	→	H_2O

Conventional Softening — Process (A)

Dealkalization by Split Stream Softening — Blending Effluents from (A) and (B)

Dealkalization by Anion Exchange — Process (C) preceded by (A)

Demineralizing — Combination of (B) and (D)

Operating costs are related principally to the cost of heat, and, to a lesser extent, the cost of utilities (cooling water, electricity, etc.) and water treating chemicals. Heat (energy) costs are relatively independent of the feed water composition.

Reverse Osmosis — involves separating water from dissolved solids by forcing the water to pass through a semi-permeable membrane which retains most of the dissolved solids. As illustrated in Fig. 18-16, this is accomplished by providing sufficient pressure on the system feedwater to overcome the normal osmotic pressure and produce a reasonable flow rate through the membrane.

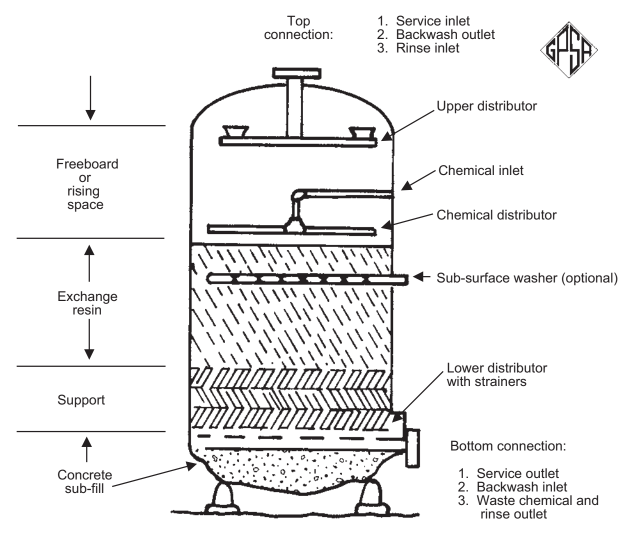
A typical brackish water with a dissolved solids content of 1500 mg/kg will have an osmotic pressure of about 100 kPa; seawater with a dissolved solids content of about 35000 mg/kg has an osmotic pressure of about 2400 kPa. The applied pressure for brackish water purification is typically in the range of 2760-4140 kPa (ga) and for seawater purification, in the range of 5500-6900 kPa (ga). Recovery of product (desalted) water with reverse osmosis units ranges from 50 to 90% of the feedwater depending upon the feedwater composition, the product water quality requirement, and the number of stages utilized.

Operating costs consist mainly of pumping costs (the pressure drop across the membrane may be from 1725 kPa to 6900 kPa, depending upon dissolved solids content and membrane selection) and membrane cleaning and replacement costs. For water containing from about 250 to 1500 mg/kg dissolved solids, an economic comparison of ion exchange and reverse osmosis is frequently necessary to select the more cost effective process. Reverse osmosis has been successfully employed for desalination of seawater. In many cases, the reverse osmosis product water must be treated by one of the ion exchange processes if high quality boiler feedwater is required.

A pretreatment system is needed to avoid fouling or excessive degradation of the membrane. Typically, pretreatment will include filtration to remove suspended particles and addition of chemicals to prevent scaling and biological growth.

FIG. 18-15

Typical Ion-Exchange Bed



Because the optimum operating temperature for reverse osmosis systems is about 24-27°C, it is frequently desirable to heat the feedwater. This represents an additional operating cost; however, because reverse osmosis is a continuous process which does not require regenerant chemicals, the cost of disposing of the waste water from the reverse osmosis system may be less than that of waste water from an ion exchange unit.

Electrodialysis — involves separating water from dissolved solids by passing the dissolved solids (ions) through a semi-permeable membrane which is relatively impervious to water. This is accomplished by means of a direct current electrical field which transports the ions through the membranes. Fig. 18-17 shows a basic electrodialysis system with alternating cation-selective and anion-selective membranes.

Recovery of product (deionized) water with electrodialysis units ranges from 50 to 90% of the feedwater depending upon the number of stages and degree of recirculation utilized. Operating costs consist mainly of power costs (typically 1.6-2.7 kwh/m^3 of product water) and membrane cleaning and replacement costs. Based upon combined capital and operating costs, the electrodialysis process is most economical when used to desalt brackish water (1000 to 5000 mg/kg dissolved solids) to a product water concentration of about 500 mg/kg dissolved solids.

A pretreatment system is usually needed to prevent fouling or degradation of the membranes. Electrodialysis units can operate over a pH of 1 to 13 and at temperatures up to about 43°C.

Deaeration (Degasifying) — Although other gases (e.g. H_2S, ammonia, methane) can be present in source or makeup water, the dissolved gases of primary concern in boiler feedwater and steam condensate are oxygen and carbon dioxide. Both are highly corrosive and should be removed to the greatest extent possible because the presence of these gases can result in significant damage to piping and equipment and the resulting corrosion products can foul boiler heat transfer surfaces. If a steam condensate treater (polisher) is utilized, high concentrations of corrosion products increase its load and

FIG. 19-1 (Cont'd)

Nomenclature

Y_{n+1} = moles of any component in the entering rich gas per mole of rich gas
Y_o = moles of any component in the gas in equilibrium with the entering lean oil, per mole of rich gas
Z = static head, m

Greek

α = relative volatility
β_{ij} = volatility factor defined in Eq 19-5
θ = correlating parameter in Eq 19-7, 19-8
σ = surface tension, dyne/cm
ρ = density, kg/m^3
ε = efficiency
μ = viscosity, $P_a \cdot s$

Subscripts

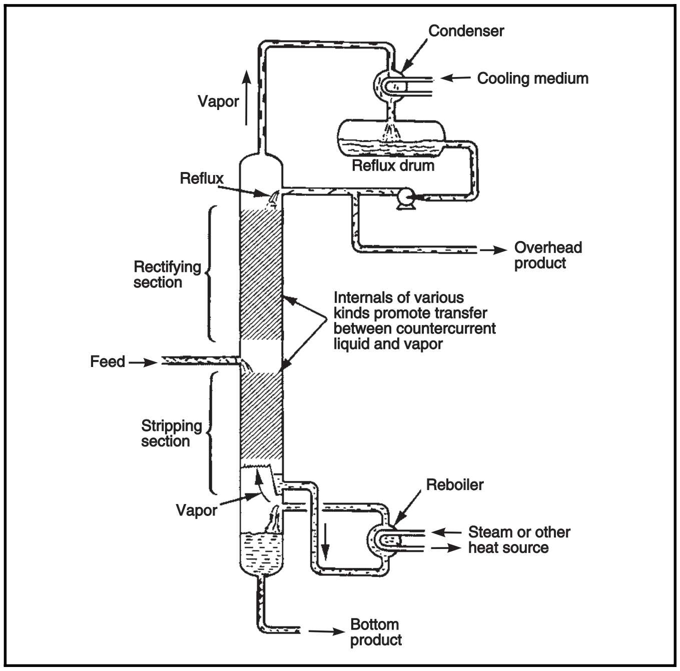
avg = average value
B = bottoms
BP = bubble point feed stream
bottom = bottom of the column
calc = calculated value
corr = corrected value
D = distillate (overhead)
F = feed
G = gas
HK = heavy key
i = any component
L = liquid
LK = light key
m = minimum
n = tray number
top = top of the column
VF = vaporized feed stream
v = vapor phase

Virtually all gas processing plants producing natural gas liquids require at least one fractionator to produce a liquid product which will meet sales specifications. The schematic of an example fractionator in Fig. 19-2 shows the various components of the system. Heat is introduced to the reboiler to produce stripping vapors. The vapor rises through the column contacting the descending liquid. The vapor leaving the top of the column enters the condenser where heat is removed by some type of cooling medium. Liquid is returned to the column as reflux to limit the loss of heavy components overhead.

Internals such as trays or packing promote the contact between the liquid and vapor streams in the column. Intimate contact of the vapor and liquid phases is required for efficient separation. Vapor entering a separation stage will be cooled which results in some condensation of heavier components. The liquid phase will be heated which results in some vaporization of the lighter components. Thus, the heavier components are concentrated in the liquid phase and eventually become the bottom product. The vapor phase is continually enriched in the light components which will make up the overhead product.

FIG. 19-2

Fractionation Schematic Diagram



The vapor leaving the top of the column may be totally or partially condensed. In a total condenser, all vapor entering the condenser is condensed to liquid and the reflux returned to the column has the same composition as the distillate or overhead product. In a partial condenser, only a portion of the vapor entering the condenser is condensed to liquid. In most partial condensers only sufficient liquid will be condensed to serve as reflux for the tower. In some cases, however, more liquid will be condensed than is required for reflux and there will actually be two overhead products, one a liquid having the same composition as the reflux and the other a vapor product which is in equilibrium with the liquid reflux.

Equilibrium Stage Concept

All calculations are performed using theoretical (equilibrium) stages. A fractionation column may be considered as a series of equilibrium flashes with two feeds and two product streams (Fig. 19-3). The vapor enters the flash from the stage below at a higher temperature and the liquid stream enters from the stage above at a lower temperature. Heat and mass transfer occur in this stage such that the exiting streams are a bubble point liquid and dew point vapor at the same temperature and pressure. The compositions of these phases are related by the equilibrium relationship of $y_i = K_i x_i$ (See Section 25). This relationship, along with heat and material balance considerations, is the basis for all fractionator design.

Types of Fractionators

The number and type of fractionators required depend on the number of products to be made and the feed composition. Typical NGL products from a fractionation process include:

- Demethanized Product (C_2+)
- Deethanized Product (C_3+)
- Ethane/Propane mixtures (EP)
- Commercial Propane
- Propane/Butane mixture (LPG)
- Butane(s)

Inhibitor losses to the hydrocarbon liquid phase are more difficult to predict. Solubility is a strong function of both the water phase and hydrocarbon phase compositions. Fig. 20-66 presents experimental data[32,33,34] showing solubility of methanol in a paraffinic hydrocarbon liquid as a function of temperature and methanol concentration. Methanol solubility in naphthenic hydrocarbons is slightly less than paraffinic, but solubility in aromatic hydrocarbons may be four to six times higher than in parafins.

Solubility of EG in the liquid hydrocarbon phase is extremely small.[29] A solubility of 40 g/m^3 of NGL is often used for design purposes. However, entrainment and other physical losses may result in total losses significantly higher than this.

Example 20-10—2.83 (10^6) Sm^3/day of natural gas leaves an offshore platform at 38°C and 8300 kPa (abs). The gas comes onshore at 4°C and 6200 kPa (abs). The hydrate temperature of the gas is 18°C. Associated condensate production is 56 m^3/Standard m^3 (10^6). The condensate has a density of 778 kg/m^3 and a molecular mass of 140. Calculate the amount of methanol and 80 mass% EG inhibitor required to prevent hydrate formation in the pipeline.

Solution Steps:

Methanol

1. Calculate the amount of water condensed per day

   from Fig. 20-4, $W_{in} = 850\ mg/Sm^3$
   $W_{in} = \underline{152\ mg/Sm^3}$
   $\Delta W = 698\ mg/Sm^3$

   Water condensed = $(2.83 \times 10^6)(698)$ = 1975 (10^6) mg/day = 1975 kg/day

2. Calculate required methanol inhibitor concentration from Eq 20-5 and 20-7 (with Fig. 20-59).

   $d = 14°C \qquad M = 32$

   Solving for X_I,

   $X_I = 0.255$, Eq 20-5

   $X_I = 0.275$, Eq 20-7 (use this value in subsequent calculations)

3. Calculate mass rate of inhibitor solution in water phase from Eq. 20-8 (assume 100% methanol is injected)

   $$m_I = \frac{X_R \bullet m_{H_2O}}{X_L - X_R} = \frac{(0.275)\,(1975)}{(1 - 0.275)} = 749 \text{ kg/day}$$

4. Estimate vaporization losses from Fig. 20-65.

   @ 4°C and 6200 kPa (abs),

   $$\text{losses} = 16.8\,(10^{-6})\ \frac{kg/m^3}{wt\%\ MeOH}$$

   daily losses = $(16.8 \times 10^{-6})(2.83 \times 10^6)(27.5)$ = 1310 kg/day

5. Estimate losses to hydrocarbon liquid phase from Fig. 20-66.

   @ 4°C and 27.5 wt% MeOH, $x_{MeOH} \approx 0.2$ mol%

   lb • mols of condensate per day –

   $$\left(\frac{2.83 \times 10^6\ Sm^3}{day}\right)\left(\frac{56m^3}{10^6\ Sm^3}\right)\left(\frac{778kg}{m^3}\right)\left(\frac{1\ kg \bullet mol}{140\ kg}\right)$$

   = 881 kg • mol/day

   kg • mol methanol = (881)(0.002) = 1.76 kg • mols/day

   kg methanol = (1.76)(32) = 56 kg/day

   Total methanol injection rate = 749 + 1310 + 56
   = 2115 kg/day

Methanol left in the gas phase can be recovered by condensation with the remaining water in downstream chilling processes. Likewise, the methanol in the condensate phase can be recovered by downstream water washing.

80 wt% EG

1. Calculate required inhibitor concentration from Eq 20-6.

   $d = 14°C \qquad M = 62$

   Solving for X_I, $X_I = 0.28$

2. Calculate mass rate of inhibitor in water phase from Eq. 20-8.

   $$m_I = \frac{(0.28)(1975)}{(0.8 - 0.28)} = 1063 \text{ kg/day}$$

Vaporization and liquid hydrocarbon losses are negligible.

Inhibitor losses represent a significant operating cost and can cause problems in downstream process units. Efficient inhibitor separation should be provided.

Low Dosage Hydrate Inhibitors (LDHIs) — LDHIs can provide significant benefits compared to thermodynamic inhibitors including:

- Significantly lower inhibitor concentrations and therefore dosage rates. Concentrations rante from 0.1 to 1.0 mass percent polymer in the free water phase, whereas alcohols can be as high as 50%
- Lower inhibitor loss caused by evaporation, particularly compared to methanol
- Reduced capital expenses through decreased chemial storage and injection rate requirements; and no need for regeneration because the chemicals are not currently recovered. These are especially appropriate for offshore where weight and space are critical to costs
- Reduced operating expenses in many cases through decreased chemical consumption and delivery frequency
- Increased production rates, where inhibitor injection capacity or flowline capacity is limited
- Lower toxicity

Kinetic Hydrate Inhibitors — KHIs were designed to inhibit hydrate formation in flowlines, pipelines, and downhole equipment operating within hydrate-forming conditions such as subsea and cold-weather environments. Their unique chemical structure significantly reduces the rate of nucleation and hydrate growth during conditions thermodynamically favorable for hydrate formation, without altering the thermodynamic hydrate formation conditions (i.e., temperature and pressure). This mechanism differs from methanol or glycol, which depress the thermodynamic hydrate formation temperature so that a flowline operates outside hydrate-forming conditions.

KHIs Compared to Methanol or Glycols— KHIs inhibit hydrate formation at a concentration range of 0.1–1.0 mass percent polymer in the free water phase. At the maximum recommended dosage, the current inhibition capabilities are –2°C of subcooling in a gas system and –7°C in an oil system with efforts continuing to expand the region of effectiveness. For relative comparison, methanol or glycol typically

FIG. 21-5

Physical Properties of Gas Treating Chemicals

	Monoethanol-amine	**Diethanol-amine**	**Triethanol-amine**	**Diglycol®-amine**	**Diisopropanol-amine**	**Selexol®**
Formula	$HOC_2H_4NH_2$	$(HOC_2H_4)_2NH$	$(HOC_2H_4)_3N$	$H(OC_2H_4)_2NH_2$	$(HOC_3H_6)_2NH$	Polyethylene glycol derivative
Molecular Wt	61.08	105.14	148.19	105.14	133.19	280
Boiling point @ 760 mm Hg, °C	170.5	269	360 (decomposes)	221	248.7	270
Freezing point, °C	10.5	28.0	22.4	–12.5	42	–28.9
Critical constants						
Pressure, kPa (abs)	5985	3273	2448	3772	3770	–
Temperature, °C	350	442.1	514.3	402.6	399.2	–
Density @ 20°C, gm/cc.	1.018	1.095	1.124	1.058 @ 15.6°C	0.999 @ 30°C	1.031 @ 25°C
Weight, kg/m^3	1016 @ 15.6°C	1089 @ 15.6°C	1123 @ 15.6°C	1057 @ 15.6°C		8.60 @ 25°C
Relative density 20°C/20°C	1.0179	1.0919 (30/20°C)	1.1258	1.0572	0.989 @ 45°C/20°C	–
Specific heat @ 15.6°C, kJ/(kg • °C)	2.55 @ 20°C	2.51	2.93	2.39	2.89 @ 30°C	2.05 @ 5°C
Thermal conductivity						
W/(m • °C) @ 20°C	0.256	0.220	–	0.209	–	0.19 @ 25°C
Latent heat of vaporization, kJ/kg	826 @ 760 mmHg	670 @ 73 mmHg	535 @ 760 mmHg	510 @ 760 mmHg	430 @ 760 mmHg	–
Heat of reaction, kJ/kg of Acid Gas						
H_2S			–930	–1568	–	–442 @ 25°C
CO_2			–1465	–1977	–	–372 @ 25°C
Viscosity, mPa • s	24.1 @ 20°C	350 @ 20°C (at 90% wt. solution)	1013 @ 20°C (at 95% wt. solution)	40 @ 16°C	870 @ 30°C 198 @ 45°C 86 @ 54°C	5.8 @ 25°C
Refractive index, N_d 20°C	1.4539	1.4776	1.4852	1.4598	1.4542 @ 45°C	–
Flash point, COC, °C	93	138	185	127	124	151

	Propylene Carbonate	**Methyldiethanol-amine**	**Sulfolane®**	**Methanol**	**10% Sodium Hydroxide**
Formula	$C_3H_6CO_3$	$(HOC_2H_4)_2NCH_3$	$C_4H_8SO_2$	CH_3OH	
Molecular Wt	102.09	119.16	120.17	32.04	19.05
Boiling point @ 760 mm Hg, °C	242	247	285	64.5	102.8
Freezing point, °C	–49.2	–23	27.6	–97.7	–10
Critical constants					
Pressure, kPa (abs)	–		5290	7956	
Temperature, °C	–		545	240	
Density @ 20°C, gm/cc.	1.2057				
Weight, kg/m^3		1040	1273 @ 30°C/30°C	792	1109
Relative density 20°C/20°C	1.203	1.0418	1.268		1.110
Specific heat @ 15.6°C, kJ/(kg • °C)	1.40	2.24	1.47 @ 30°C	2.47 @ 5°-10°C	3.76
Thermal conductivity					
W/(m • °C) @ 20°C	0.21 @ 10°C	0.275	0.197 @ 38°C	0.215	
Latent heat of vaporization, kJ/kg	484 @ 760 mmHg	476	525 @ 100°C	1103 @ 760 mmHg	
Heat of reaction, kJ/kg of Acid Gas					
H_2S	–				
CO_2					
Viscosity, mPa • s	1.67 @ 38°C	1.01 @ 20°C 33.8 @ 40°C	10.3 @ 30°C 6.1 @ 50°C 2.5 @ 100°C 1.4 @ 150°C 0.97 @ 200°C	0.6 @ 20°C	1.83 @ 20°C 0.97 @ 50°C 0.40 @ 100°C
Refractive index, N_d 20°C	1.4209	1.469	1.481 @ 30°C	1.3286	
Flash point, COC, °C	132.2	129.4	177	14	

MERCURY REMOVAL

Calgon

Solid adsorbents can remove mercury from gas to produce residuals in the range of 0.01–0.001 µg/Nm^3. Calgon sulfur impregnated HGR® (4 x 10 mesh) and HGR®-P (4 mm dia.) carbons are used for mercury removal and indicate designs removing mercury down to very low levels. Removal of both inorganic and organic mercury is achieved. By first drying the gas the degree of mercury removal increases. The sulfur impregnate reacts with the mercury to produce a mercury sulfide that is fixed in the carbon microstructure.

Merespec™

Johnson Matthey Catalysts supplies MereSpec™ fixed bed absorbents for removal of traces, elemental and organic, of mercury from hydrocarbon liquids and gases. The absorbents have been shown to be capable of providing the outlet mercury concentration normally specified for LNG production and are in service in several European locations including an offshore oil/gas production platform. Merespec™ is a trademark of Synetix.

Desorex

Activated carbon provides only a limited storage capacity for the strictly physical adsorption of mercury. Desorex HGD2S and HGD4S from Donau Carbon can be employed to bind mercury through the process of chemical adsorption involving oxidation and adsorption in the form of stable compound or fixation in metallic form as an amalgam. Many reference installations of these Desorex products for the purification of natural gases to levels as low as 10 µg/m^3 of mercury have been realized over a long period of time.

HgSIV

UOP supplies HgSIV adsorbents which are molecular sieves coated with elemental silver. Mercury in the gas is trapped by amalgamation with the silver. The adsorbent also serves the dual function of dehydrating the gas. HgSIV is regenerated thermally, just like molecular sieves for dehydration. This material can be added as a layer to existing molecular sieve dryers.[58] However, one must take care to appropriately handle the regeneration gas in this case, as it will contain mercury.

CMG 275 and 273

Another mercury trapping material labeled CMG 275 was developed by IFP and Procatalyse (today Axens) which is sulfur supported onto mesoporous alumina. The advantage of mesoporous alumina based product is its resistance to capillary condensation.[59] The larger pore size of this material, compared to carbon based trapping materials permits utilization under near dew point conditions. In another material from Axens labeled CMG 273 the trapping component is anchored on the alumina carrier making it completely insoluble in liquid hydrocarbons and water. The material has been subjected at gas plant sites to both DEA and liquid hydrocarbon carry-over with no active phase leaching. This same material has been used to eliminate elemental mercury from LPG and full range condensates.

Organic Mercury Removal

Removal of all forms of organic mercury compounds from natural gases and liquids requires firstly the conversion of the compounds to elemental metallic mercury followed by trapping materials to remove the metallic mercury formed. This requires in the first stage some hydrogen for the organo-mercury hydrogenolysis with a suitable catalyst. The first stage catalyst such as MEP 841 also traps arsenic and lead impurities in the feed. The two stage impurities removal process is called RAM and is available from Axens.

ACID GAS INJECTION

In some cases, it is possible to compress the acid gas as generated from a chemical or a physical solvent process and inject it into a depleted, non-producing, or even a producing reservoir[60]. A key consideration is the phase behavior of the acid gas mixture. Depending on temperature, pressure, and composition, the acid gas may be injected as a liquid or gas, or as a dense phase. Dehydration is usually necessary to avoid corrosion or hydrate formation. However, in many cases, a minimum in water-holding capacity occurs with respect to pressure. Thus, it may be possible to compress the gas to a given level, cool it, and drop out the liquid water. Further compression increases the capacity of acid gas to hold water, so that water drop out should not occur in the downstream piping or well. The phase behavior of acid gas mixtures is complex, and careful consideration must be given to the design of the project.[61,62]

There have been a number of successful acid gas injection projects, with sulfur contents of 1 to more than 70 tons per day.[63]

LIQUID HYDROCARBON TREATING

As a guide in the selection of the method of treating to be used, the following characteristics of each are given:

Regenerated Caustic

1. Can handle large volumes of hydrocarbon.
2. Primarily for removing methyl and ethyl mercaptans.
3. Capable of producing a doctor-sweet product.
4. Reduces the total sulfur content of treated product.

Perco Solid Copper Chloride

1. Can treat gasoline streams with relatively high mercaptan content.
2. Suited for small flow rate.
3. Sulfur content not reduced.
4. Water content must be only that of saturation.
5. Hydrogen sulfide must be removed ahead of contact with bed.
6. Excess regeneration oxygen may cause corrosion downstream of bed. Gasoline with components that may be affected by oxygen, such as olefins, should not be treated with this process.
7. Capable of producing doctor-sweet product.

Batch Caustic Wash

1. Can use a single wash.
2. Best suited for streams with low mercaptan content (if mercaptan removal is important).
3. Primarily for removing trace amounts of hydrogen sulfide and methyl and ethyl mercaptans.
4. Disposing of spent caustic can be a major consideration.

8. Correct the pseudo-density to the actual pressure using Fig. 23-15. Add the correction to the pseudo-density from Step 7.
9. Correct the density at 15°C and pressure to the actual temperature using Fig. 23-17. Add the correction to the density from Step 8.

This procedure is not valid in the critical region. Mixtures at temperatures greater than 65°C that contain more than 60 mol% methane or more than 80 mol% CO_2 are problem areas. Outside the near-critical region, calculated densities usually are within 5% of experimental data[35] and errors are rarely greater than 8%. The best accuracy occurs for mixtures containing mostly C_5+ with relatively small amounts of dissolved gaseous components (errors are usually less than 3%). Note that densities of C_2+, C_3+, CO_2+, or C_4+ mixtures can be calculated by this procedure at various temperatures and pressures, and that the gaseous components need not be present.

Example 23-3 — Fig. 23-16 illustrates the procedure outlined above to calculate necessary numbers for the following example.

$$\text{Density of } C_3 \text{ plus} = \frac{\text{Wt of } C_3 \text{ plus}}{\text{Vol of } C_3 \text{ plus}} = \frac{44.836\,\text{kg}}{0.0632\,\text{m}^3}$$

$$= 709.5\,\text{kg/m}^3$$

$$\text{Mass \% } C_2 \text{ in } C_2 \text{ plus} = \frac{0.567}{0.567 + 44.836} \bullet 100 = 1.25\%$$

Density of C_2 plus from Fig. 23-14 = 705 kg/m^3

$$\text{Density of } CO_2 \text{ plus} = \frac{62.889\text{ kg}}{0.0845\text{ m}^3} = 744.52\,\text{kg/m}^3$$

$$\text{Mass \% } CH_4 \text{ in Total} = \frac{3.352}{66.241} \bullet 100 = 5.1\%$$

Pseudodensity of mixture at 15°C and 101.325 kPa (abs) from Fig. 23-14 = 680 kg/m^3

Pressure correction to 12 000 kPa (abs) from Fig. 23-15 = 12.5

Density at 15°C and 12 000 kPa (abs) = 680 + 12.5 = 692.5 kg/m^3

Temperature correction to 50°C from Fig. 23-17 = –30

Density at 50°C and 12 000 kPa (abs) = 692.5 – 30

= 662.5 kg/m^3

(Density by EZ*THERMO version of SRK using Costald[92] 668 kg/m^3.

Experimental density[35] at 50°C and 12 000 kPa (abs) = 660 kg/m^3

Error = (662.5 – 660)/660 = 0.0038 = 0.38%

FIG. 23-18

Characterizing Boiling Points of Petroleum Fractions (From API Technical Data Book)

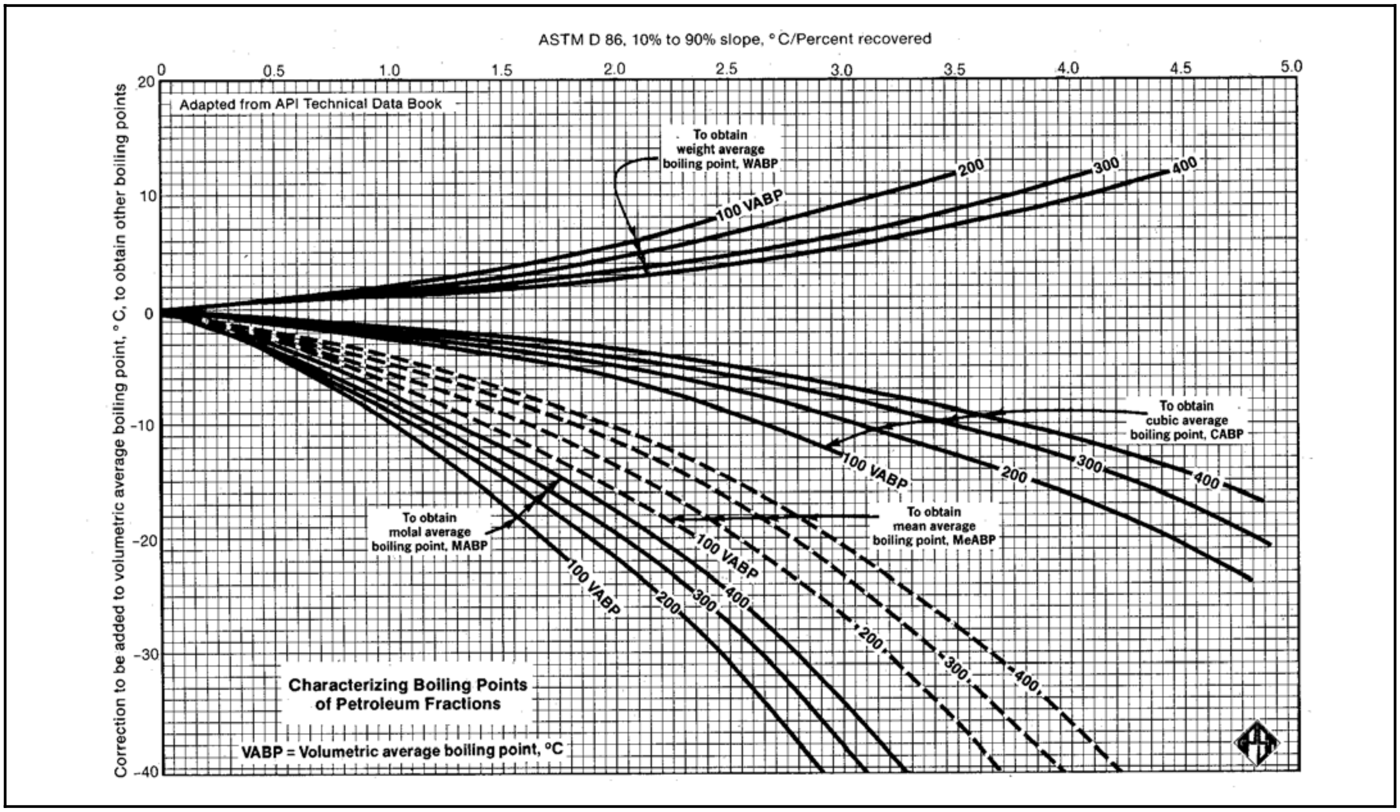
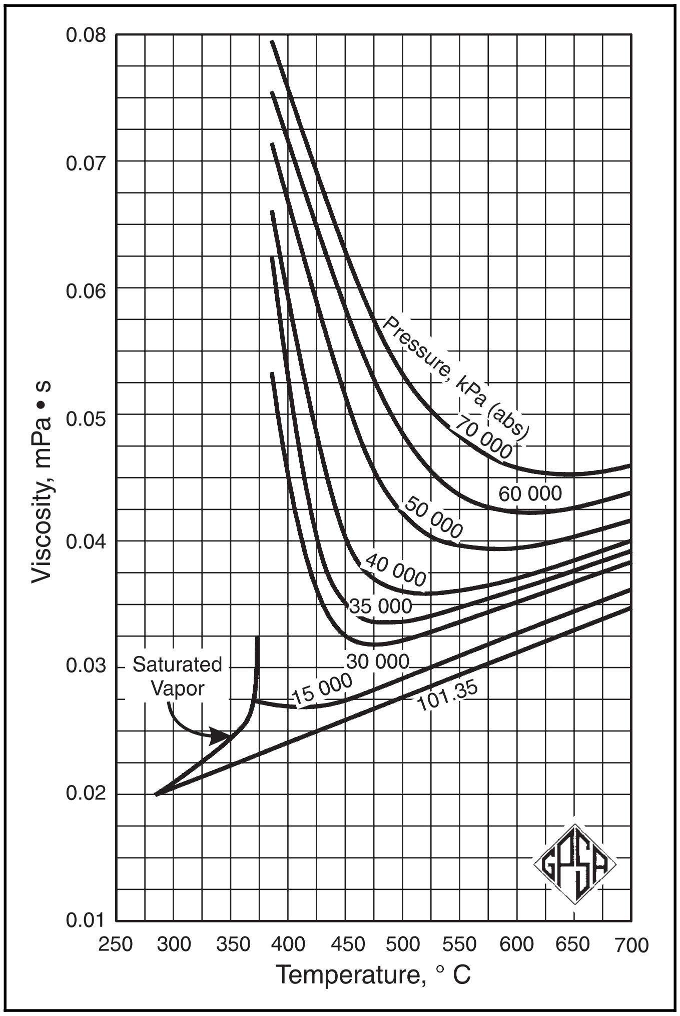


FIG. 23-29

Viscosity of Steam[46, 47]



$$K_w = \frac{\sqrt[3]{(163 + 273) \bullet 1.8}}{0.7688} = 11.99$$

$$\ln (B) = 4.717 + (0.00526)(163 + 273) = 7.01$$

$$B = 1108$$

$$A = \left[91.83\,(436)^{-0.175} - 29.263\right]\frac{11.99}{1108} = 0.02638$$

The same constants are employed at 40°C and 100°C.

$$\eta = 0.02638 \bullet e^{\frac{1108}{313}} = 0.909 \text{ cs at } 40°C$$

$$\eta = 0.02638 \bullet e^{\frac{1108}{373}} = 0.513 \text{ cs at } 100°C$$

The reported values are 0.93 and 0.52 centistokes, respectively.

THERMAL CONDUCTIVITY

Thermal conductivity for natural gas mixtures at elevated pressure can be calculated from an atmospheric value and a pressure correction. Figs. 23-31 through 23-36 present low pressure thermal conductivity data for gases developed from published data[51, 54]. The pressure correction of Lenoir *et al.*[52] shown in Fig. 23-32 applies to these low pressure data. The thermal conductivity of liquid paraffin hydrocarbons appears in Fig. 23-35 and the thermal conductivity of liquid petroleum fractions appears in Fig. 23-36.

Example 23-11 — Find the thermal conductivity of a 25 molecular mass natural gas at 4826 kPa (abs) and 140°C. $T_c = 244$ K, $P_c = 4550$ k Pa (abs)

Solution Steps

From Fig. 23-31, at 140°C:

$$k_A = 0.0445 \text{ W/(m} \bullet °C)$$

$$T_r = (140 + 273)/244 = 1.69$$

$$P_r = 4826/4550 = 1.06$$

From Fig. 23-32:

$$k/k_A = 1.13$$

$$k = (1.13)(0.0445) = 0.0503 \text{ W/(m} \bullet °C)$$

Stiel and Thodos[53] present another method for estimating thermal conductivity. To determine the thermal conductivity of a gaseous mixture of defined components, the thermal conductivity of each component at the given temperature is read from the charts provided and the thermal conductivity of the mixture is determined by the "cube root rule[56]." This rule is applicable to mixtures of simple gases; it does not apply to mixtures containing CO_2 because its thermal conductivity goes through a maximum.

FIG. 23-30

Calculation of Viscosity of a Gas Mixture

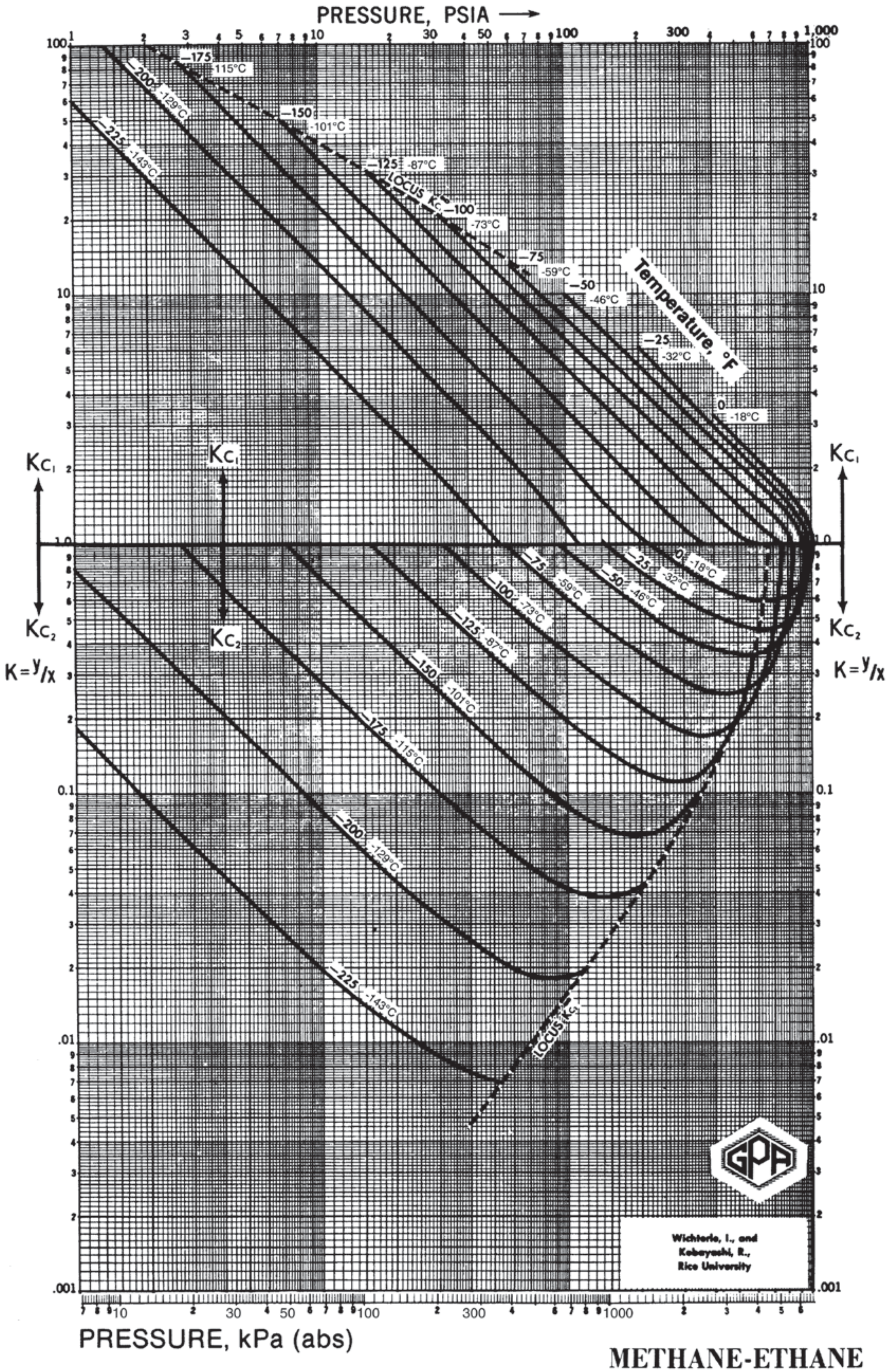
	Mole Fraction, y_i	Molecular Mass, MWi	P_{ci}, kPa (abs)	T_{ci}, K	V_{ci}, m^3/kg	$V_{ci} = MW_i \bullet v_{ci}$ m^3/kg mole	$Z_{ci} = \frac{P_{ci} \bullet V_{ci}}{8.31 \bullet T_{ci}}$
CH_4	0.80	16.043	4599	190.56	0.00617	0.0990	0.2875
N_2	0.15	28.013	3398	126.21	0.00318	0.0891	0.2887
CO_2	0.05	44.010	7374	304.11	0.00214	0.0942	0.2749
Mixture	1.00	19.237	—	186.59	—	0.0973	0.2871
	Σ	$MW_{cm} = \Sigma y_i \bullet MW_i$		$T_{cm} = \Sigma y_i \bullet T_{ci}$		$V_{cm} = \Sigma y_i \bullet V_{ci}$	$Z_{cm} = \Sigma y_i \bullet Z_{ci}$

FIG. 24-37

Saturated Steam: Pressure Table

Symbols Used in Steam Tables: P—pressure, kPa (abs); T—thermodynamic temperature, °C; v—specific volume, m^3/kg; u—specific internal energy, kJ/kg; h—specific enthalpy, kJ/kg s—specific entropy, kJ/(kg•K) Adapted from Joseph H. Keenan, Frederick G. Keyes, Philip G. Hill, and Joan G. Moore, Steam Tables, (New York: John Wiley & Sons, Inc., 1969).

		Specific Volume		**Internal Energy**			**Enthalpy**			**Entropy**		
Press. kPa P	**Temp. °C** T	**Sat. Liquid** v_f	**Sat. Vapor** v_g	**Sat. Liquid** u_f	**Evap.** u_{fg}	**Sat. Vapor** u_{fg}	**Sat. Liquid** h_{fg}	**Evap.** h_{fg}	**Sat. Vapor** h_g	**Sat. Liquid** s_f	**Evap.** s_{fg}	Sat. Vapor s_g
0.6113	0.01	0.001 000	206.14	00	2375.3	2375.3	01.	2501.3	2501.4	.0000	9.1562	9.1562
1.0	6.98	0.001 000	129.21	29.30	2355.7	2385.0	29.30	2484.9	2514.2	.1059	8.8697	8.9756
1.5	13.03	0.001 001	87.98	54.71	2338.6	2393.3	54.71	2470.6	2525.3	.1957	8.6322	8.8279
2.0	17.50	0.001 001	67.00	73.48	2326.0	2399.5	73.48	2460.0	2533.5	.2607	8.4629	8.7237
2.5	21.08	0.001 002	54.25	88.48	2315.9	2404.4	88.49	2451.6	2540.0	.3120	8.3311	8.6432
3.0	24.08	0.001 003	45.67	101.04	2307.5	2408.5	101.05	2444.5	2545.5	.3545	8.2231	8.5776
4.0	28.96	0.001 004	34.80	121.45	2293.7	2415.2	121.46	2432.9	2554.4	.4226	8.0520	8.4746
5.0	32.88	0.001 005	28.19	137.81	2282.7	2420.5	137.82	2423.7	2561.5	.4764	7.9187	8.3951
7.5	40.29	0.001 008	19.24	168.78	2261.7	2430.5	168.79	2406.0	2574.8	.5764	7.6750	8.2515
10	45.81	0.001 010	14.67	191.82	2246.1	2437.9	191.83	2392.8	2584.7	.6493	7.5009	8.1502
15	53.97	0.001 014	10.02	225.92	2222.8	2448.7	225.94	2373.1	2599.1	.7549	7.2536	8.0085
20	60.05	0.001 017	7.649	251.38	2205.4	2456.7	251.40	2358.3	2609.7	.8320	7.0766	7.9085
25	64.97	0.001 020	6.204	271.90	2191.2	2463.1	271.93	2346.3	2618.2	.8931	6.9383	7.8314
30	69.10	0.001 022	5.229	289.20	2179.2	2468.4	289.23	2336.1	2625.3	.9439	6.8247	7.7686
40	75.87	0.001 027	3.993	317.53	2159.5	2477.0	317.58	2319.2	2636.8	1.0259	6.6441	7.6700
50	81.33	0.001 030	3.240	340.44	2143.4	2483.9	340.49	2305.4	2645.9	1.0910	6.5029	7.5939
75	91.78	0.001 037	2.217	384.31	2112.4	2496.7	384.39	2278.6	2663.0	1.2130	6.2434	7.4564
100	99.63	0.001 043	1.6940	417.36	2088.7	2506.1	417.46	2258.0	2675.5	1.3026	6.0368	7.3594
125	105.99	0.001 048	1.3749	444.19	2069.3	2513.5	444.32	2241.0	2685.4	1.3740	5.9104	7.2844
150	111.37	0.001 053	1.1593	466.94	2052.7	2519.7	467.11	2226.5	2693.6	1.4336	5.7897	7.2233
175	116.06	0.001 057	1.0036	486.80	2038.1	2524.9	486.99	2213.6	2700.6	1.4849	5.6868	7.1717
200	120.23	0.001 061	0.8857	504.49	2025.0	2529.5	504.70	2201.9	2706.7	1.5301	5.5970	7.1271
225	124.00	0.001 064	0.7933	520.47	2013.1	2533.6	520.72	2191.3	2712.1	1.5706	5.5173	7.0878
250	127.44	0.001 067	0.7187	535.10	2002.1	2537.2	535.37	2181.5	2716.9	1.6072	5.4455	7.0527
275	130.60	0.001 070	0.6573	548.59	1991.9	2540.5	548.89	2172.4	2721.3	1.6408	5.3801	7.0209
300	133.55	0.001 073	0.6058	561.15	1982.4	2543.6	561.47	2163.8	2725.3	1.6718	5.3201	6.9919
325	136.30	0.001 076	0.5620	572.90	1973.5	2546.4	573.25	2155.8	2729.0	1.7006	5.2646	6.9652
350	138.88	0.001 079	0.5243	583.95	1965.0	2548.9	584.33	2148.1	2732.4	1.7275	5.2130	6.9405
375	141.32	0.001 081	0.4914	594.40	1956.9	2551.3	594.81	2140.8	2735.6	1.7528	5.1647	6.9175
400	143.63	0.001 084	0.4625	604.31	1949.3	2553.6	604.74	2133.8	2738.6	1.7766	5.1193	6.8959
450	147.93	0.001 088	0.4140	622.77	1934.9	2557.6	623.25	2120.7	2743.9	1.8207	5.0359	6.8565
500	151.86	0.001 093	0.3749	639.68	1921.6	2561.2	640.23	2108.5	2748.7	1.8607	4.9606	6.8213
550	155.48	0.001 097	0.3427	655.32	1909.2	2564.5	655.93	2097.0	2753.0	1.8973	4.8920	6.7893
600	158.85	0.001 101	0.3157	669.90	1897.5	2567.4	670.56	2086.3	2756.8	1.9312	4.8288	6.7600
650	162.01	0.001 104	0.2927	683.56	1886.5	2570.1	684.28	2076.0	2760.3	1.9627	4.7703	6.7331
700	164.97	0.001 108	0.2729	696.44	1876.1	2572.5	697.22	2066.3	2763.5	1.9922	4.7158	6.7080
750	167.78	0.001 112	0.2556	708.64	1866.1	2574.7	709.47	2057.0	2766.4	2.0200	4.6647	6.6847
800	170.43	0.001 115	0.2404	720.22	1856.6	2576.8	721.11	2048.0	2769.1	2.0462	4.6166	6.6628
850	172.96	0.001 118	0.2270	731.27	1847.4	2578.7	732.22	2039.4	2771.6	2.0710	4.5711	6.6421
900	175.38	0.001 121	0.2150	741.83	1838.6	2580.5	742.83	2031.1	2773.9	2.0946	4.5280	6.6226
950	177.69	0.001 124	0.2042	751.95	1830.2	2582.1	753.02	2023.1	2776.1	2.1172	4.4869	6.6041
1000	179.91	0.001 127	0.194 44	761.68	1822.0	2583.6	762.81	2015.3	2778.1	2.1387	4.4478	6.5865
1100	184.09	0.001 133	0.177 53	780.09	1806.3	2586.4	781.34	2000.4	2781.7	2.1792	4.3744	6.5536
1200	187.99	0.001 139	0.163 33	797.29	1791.5	2588.8	798.65	1986.2	2784.8	2.2166	4.3067	6.5233
1300	191.64	0.001 144	0.151 25	813.44	1777.5	2591.0	814.93	1972.7	2787.6	2.2515	4.2438	6.4953
1400	195.07	0.001 149	0.140 84	828.70	1764.1	2592.8	830.30	1959.7	2790.0	2.2842	4.1850	6.4693
1500	198.32	0.001 154	0.131 77	843.16	1751.3	2594.5	844.89	1947.3	2792.2	2.3150	4.1298	6.4448
1750	205.76	0.001 166	0.113 49	876.46	1721.4	2597.8	878.50	1917.9	2796.4	2.3851	4.0044	6.3896
2000	212.42	0.001 177	0.099 63	906.44	1693.8	2600.3	908.79	1890.7	2799.5	2.4474	3.8935	6.3409
2250	218.45	0.001 187	0.088 75	933.83	1668.2	2602.0	936.49	1865.2	2801.7	2.5035	3.7937	6.2972
2500	223.99	0.001 197	0.079 98	959.11	1644.0	2603.1	962.11	1841.0	2803.1	2.5547	3.7028	6.2575
3000	233.90	0.001 217	0.066 68	1004.78	1599.3	2604.1	1008.42	1795.7	2804.2	2.6457	3.5412	6.1869
3500	242.60	0.001 235	0.057 07	1045.43	1558.3	2603.7	1049.75	1753.7	2804.2	2.7253	3.4000	6.1253
4000	250.40	0.001 252	0.049 78	1082.31	1520.0	2602.3	1087.31	1714.1	2801.4	2.7964	3.2737	6.0701
5000	263.99	0.001 286	0.039 44	1147.81	1449.3	2597.1	1154.23	1640.1	2794.3	3.9202	3.0532	5.9734
6000	275.64	0.001 319	0.032 44	1205.44	1384.3	2589.7	1213.35	1571.0	2784.3	3.0267	2.8625	5.8892
7000	285.88	0.001 351	0.027 37	1257.55	1323.0	2580.5	1267.00	1505.1	2772.1	3.1211	2.6922	5.8133
8000	295.06	0.001 384	0.023 52	1305.57	1264.2	2569.8	1316.64	1441.3	2758.0	3.2068	2.5364	5.7432
9000	303.40	0.001 418	0.020 48	1350.51	1207.3	2557.8	1363.26	1378.9	2742.1	3.2858	2.3915	5.6772
10000	311.06	0.001 452	0.018 026	1393.04	1151.4	2544.4	1407.56	1317.1	2724.7	3.3596	2.2544	5.6141
11000	318.15	0.001 489	0.015 987	1433.7	1096.0	2529.8	1450.1	1255.5	2705.6	3.4295	2.1233	5.5527
12000	324.75	0.001 527	0.014 263	1473.0	1040.7	2513.7	1491.3	1193.6	2684.9	3.4962	1.9962	5.4924
13000	330.93	0.001 567	0.012 780	1511.1	985.0	2496.1	1531.5	1130.7	2662.2	3.5606	1.8718	5.4323
14000	336.75	0.001 611	0.011 485	1548.6	928.2	2476.8	1571.1	1066.5	2637.6	3.6232	1.7485	5.3717
15000	342.24	0.001 658	0.010 337	1585.0	869.8	2455.5	1610.5	1000.0	2610.5	3.6848	1.6249	5.3098
16000	347.44	0.001 711	0.009 306	1622.7	809.0	2431.7	1650.1	930.6	2580.6	3.7461	1.4994	5.2455
17000	352.37	0.001 770	0.008 364	1660.2	744.8	2405.0	1690.3	856.9	2547.2	3.8079	1.3698	5.1777
18000	357.06	0.001 840	0.007 489	1698.9	675.4	2374.3	1732.0	777.1	2509.1	3.8715	1.2329	5.1044
19000	361.54	0.001 924	0.006 657	1739.9	598.1	2338.1	1776.5	688.0	2464.5	3.9388	1.0839	5.0228
20000	365.81	0.002 036	0.005 834	1785.6	507.5	2293.0	1826.3	583.4	2409.7	4.0139	.9130	4.9269
21000	369.89	0.002 207	0.004 952	1842.1	388.5	2230.6	1888.4	446.2	2334.6	4.1075	.6938	4.8013
22000	373.80	0.002 742	0.003 568	1961.9	125.2	2087.1	2022.2	143.4	2165.6	4.3110	.2216	5.5327
22090	374.14	0.003 155	0.003 155	2029.6	0	2029.6	2099.3	0	2099.3	4.4298	0	4.4298



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