

SECTION 19

Fractionation and Absorption

Fractionation

Fractionation is a unit operation utilized to separate mixtures into individual products. Fractionation involves separating components by relative volatility (a). The difficulty

of a separation is directly related to the relative volatility of the components and the required purity of the product streams.

FIG. 19-1
Nomenclature

a'_t = tube flow area, m^2	NP = number of passes in a tray
a_t = total tube flow area, m^2	N_t = number of tubes
A = absorption factor used in Equation 19-28	ΔP = pressure drop, kPa
A_c = cross sectional area, m^2	q = moles of saturated liquid in the feed per mole of feed
A_t = heat transfer surface, m^2	Q = heat transfer duty, kw
AAM = tray active area, m^2	Q_c = condenser duty, kw
ADM = tray downcomer area, m^2	R = reflux ratio, moles of reflux divided by moles of net overhead product
ATM = tower cross sectional area, m^2	Re = Reynolds number, dimensionless
b = exponent in Equation 19-5 and 19-6	s = relative density
B = bottoms product rate, moles/unit time	S = number of stages
C = coefficient in Equation 19-11, m/h	S_T = stripping factor used in Equation 19-31
CAF = vapor capacity factor, corrected, m/s	S_F = separation factor defined by Equation 19-1
CAF_o = vapor capacity factor, uncorrected, m/s	TS = tray spacing, mm
CFS = vapor loading, m^3/s	U_D = overall heat transfer coefficient, $W/(m^2 \cdot ^\circ C)$
D' = diameter, m	v = specific volume, m^3/kg
D = distillate (overhead) product rate, moles/unit time	v_{max} = maximum velocity, m/h
D_T = tower diameter, m	v_i = specific volume of the inlet, m^3/kg
E_a = absorption efficiency, Equation 19-30	v_o = specific volume of the outlet, m^3/kg
E_s = stripping efficiency, Equation 19-32	V = vapor rate, moles/unit time
f = friction factor (Moody friction factor/100), m^2/dm^2	V_1 = vapor rate leaving top tray, moles/unit time
F = feed rate, moles/unit time	V_{max} = volumetric vapor flow rate, m^3/h
F_p = packing factor	VD_{dsg}^* = downcomer velocity (uncorrected), $m^3/h/m^2$
FF = flooding factor used in Equation 19-17	VD_{dsg} = downcomer velocity (corrected), $m^3/h/m^2$
FPL = flow path length, m	V_{load} = vapor loading defined by Equation 19-13, m^3/s
G_t = mass velocity, $kg/(h \cdot m^2)$	V_o = stripping medium rate, moles/unit time
G_p = tower vapor loading, $kg/(m^2 \cdot s)$	w = weight flow, kg/h
GPM = tower liquid loading, m^3/min	x = liquid mole fraction
H = enthalpy, kJ/kg	X = liquid rate, moles/unit time
HETP = height of packing equivalent to a theoretical plate	X_{m+1} = moles of a component in the rich oil entering a stripper per mole of rich oil entering the stripper
HTU = height of a transfer unit	x_1 = moles of a component in the lean oil per mole of rich oil
K = equilibrium K-value, y/x	X_o = moles of a component in the liquid in equilibrium with the stripping medium per mole of entering rich oil
L_o = liquid reflux rate, moles/unit time	y = vapor mole fraction
L_p = liquid loading, $kg/(m^2 \cdot sec)$	Y_i = moles of any component in the lean gas leaving the absorber per mole of rich gas
L = liquid rate, moles/unit time	Y_{n+1} = moles of any component in the entering rich gas per mole of rich gas
L_t = tube length, m	
L_{m+1} = rich oil entering the stripper, moles/unit time	
m = number of stripping stages	
M = mass flow rate, kg/h	
n = number of absorber stages	
N_m = minimum number of theoretical stages	

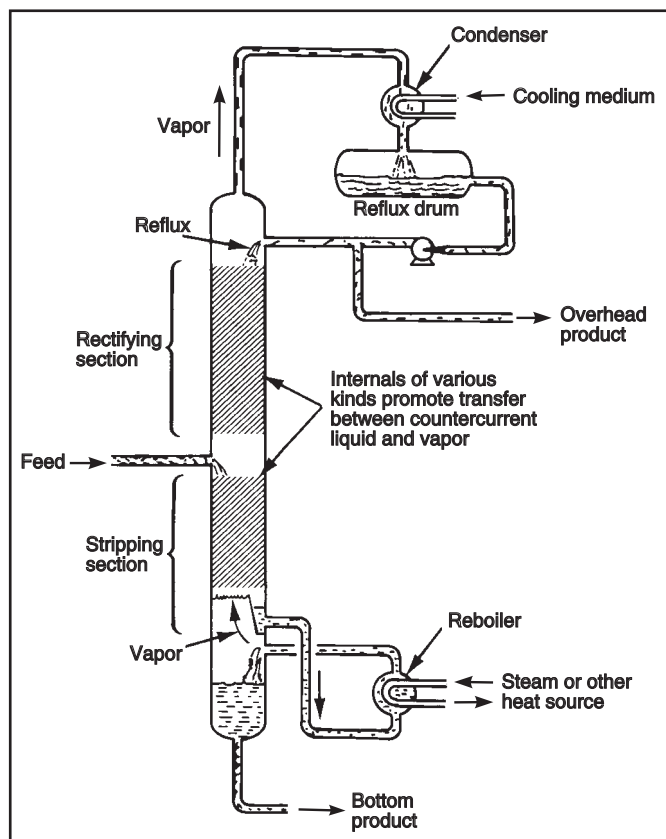
FIG. 19-1 (Cont'd)
Nomenclature

Y_o = moles of any component in the gas in equilibrium with the entering lean oil, per mole of rich gas	BP = bubble point feed stream
Z = static head, m	bottom = bottom of the column
Greek	calc = calculated value
α = relative volatility	corr = corrected value
β_{ij} = volatility factor defined in Equation 19-5	D = distillate (overhead)
θ = correlating parameter in Equation 19-7, 19-8	F = feed
σ = surface tension, dyne/cm	G = gas
ρ = density, kg/m ³	HK = heavy key
ε = efficiency	i = any component
μ = viscosity, $P_a \cdot s$	L = liquid
Subscripts	LK = light key
avg = average value	m = minimum
B = bottoms	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 pro-

duce 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.

FIG. 19-2
Fractionation Schematic Diagram



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.

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)
- Butane/Gasoline mixtures
- Natural Gasoline
- Mixtures with a vapor pressure specification

See Section 1 for definitions and Section 2 for product specifications.

An example fractionation train used to produce three products is illustrated in Fig. 19-4. The feed stream contains too much ethane to be included in the products; thus, the first

FIG. 19-3
Basic Fractionation Model

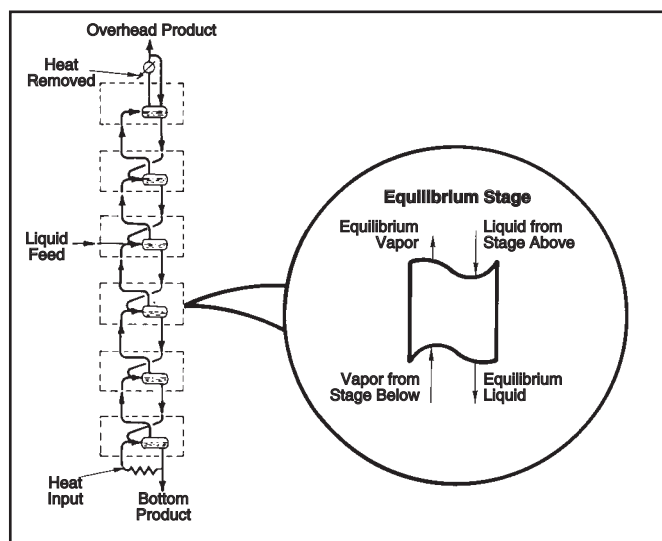
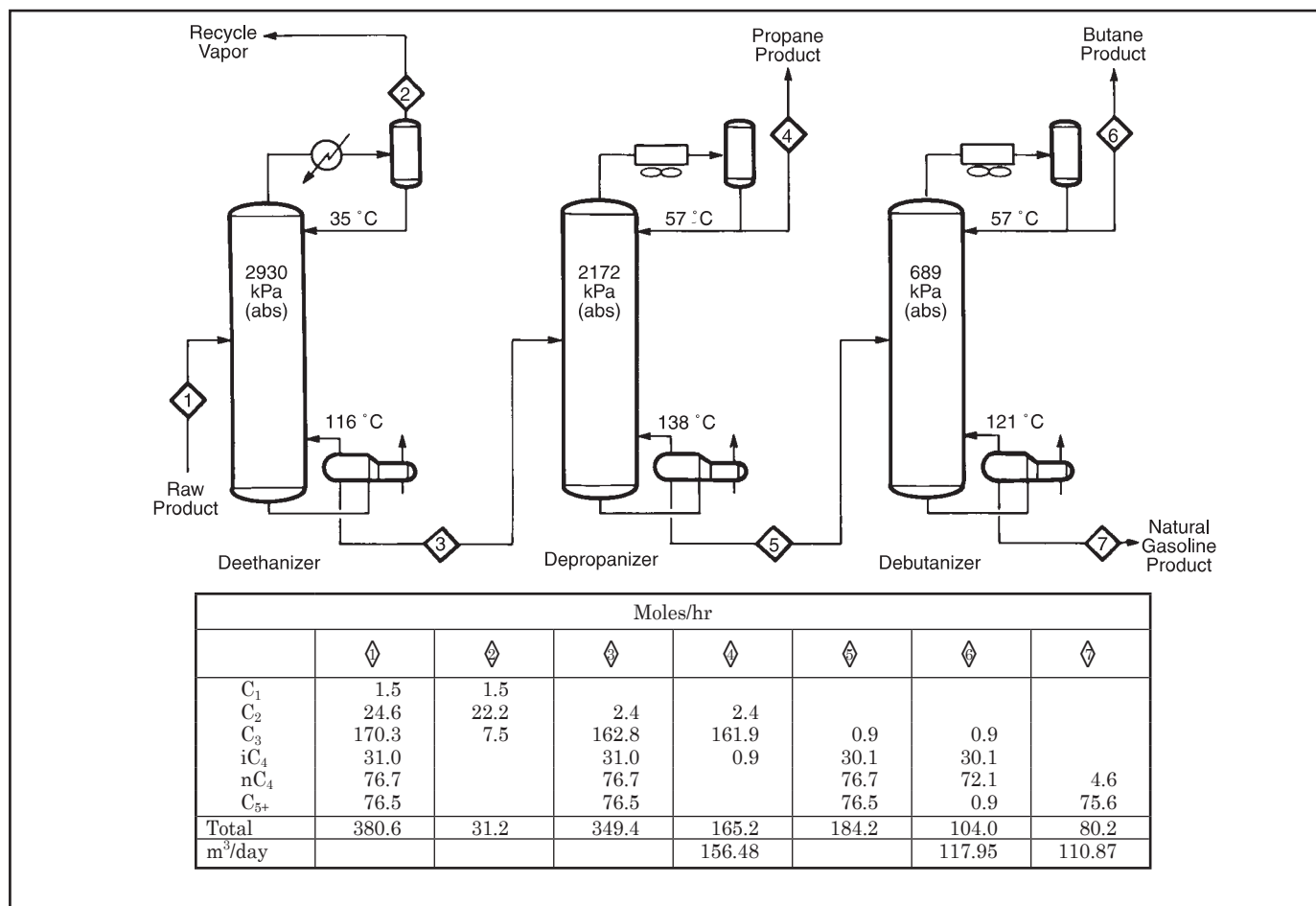


FIG. 19-4
Fractionation Train

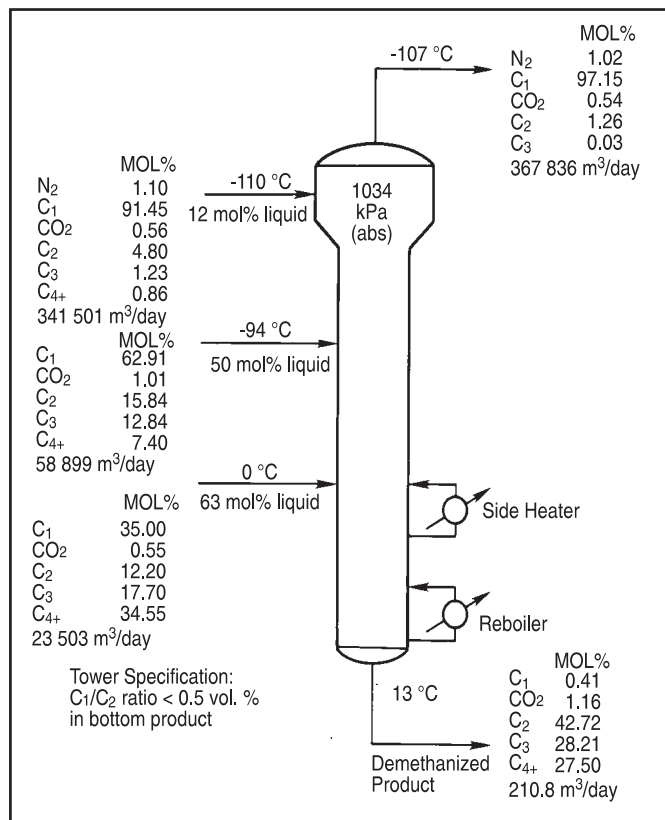


column is a deethanizer. The overhead stream is recycled to the upstream processing plant or sent to a fuel system. The bottom product from this column could be marketed as a deethanized product. The second column, a depropanizer, produces a specification propane product overhead. The bottom product, a butane-gasoline mixture, is often sold to a pipeline without further processing. The third column, a debutanizer, separates the butane and gasoline products. This separation is controlled to limit the vapor pressure of the gasoline. The overhead butane product can be sold as a mixture or an additional column can be used to separate the iso-butane and normal-butane.

Another class of fractionators uses no external reflux condenser to produce liquid for contact with the fractionator vapor. One such tower is a demethanizer as is found in cryogenic plants (Fig. 19-5). The top feed being 12 mol % liquid at low temperature provides the reflux. This liquid along with the other feeds provides the liquid loading for this tower. The reboiler is the control point for the bottom product purity. The overhead composition is a function of the upstream process units.

Applications also exist for a fractionator with a top liquid feed. These are commonly used for crude stabilizers or deethanizers. As with the demethanizer, this column produces a specification bottom product and an overhead stream whose composition is determined by upstream process units. This is an economical approach to producing a single product but is limited in separation efficiency. Better recovery or sharper separation can be achieved by adding a reflux condenser and rectifying section.

FIG. 19-5
Demethanizer Example



PRODUCT SPECIFICATIONS

A material balance around the column is the first step in fractionation calculations. In order to perform this balance, assumption of the product stream compositions must be made. There are three ways of specifying a desired product from a fractionator:

- A percentage recovery of a component in the overhead or bottom stream.
- A composition of one component in either product.
- A specific physical property, such as vapor pressure, for either product.

The recovery and composition specifications can be used directly in the material balance. However, property specifications are used indirectly. For instance if vapor pressure is a desired specification of a product, a material balance is performed with an assumed component split. The calculated vapor pressure of the resulting stream is then compared with the desired value and the material balance redone until reasonable agreement is reached.

In a multicomponent mixture, there are typically two components which are the “keys” to the separation. The light key component is defined as the lightest component in the bottom product in a significant amount. The heavy key component is the heaviest component in the overhead product in a significant amount. Normally, these two components are adjacent to each other in the volatility listing of the components. For hand calculations, it is normally assumed for material balance purposes that all components lighter than the light key are produced overhead and all components heavier than the heavy key are produced with the bottom product. By definition, the key components will be distributed between the product streams.

Example 19-1 — For the given feed stream, estimate the product stream compositions for 98% propane recovered in the overhead product with a maximum iso-butane content of the overhead stream of 1%.

Feed:	C ₂	2.4
	C ₃	162.8
	iC ₄	31.0
	nC ₄	76.7
	C ₅	76.5
		349.4 moles

Solution Steps

For Propane (light key):

$$\text{Moles in overhead} = (0.98) 162.8 = 159.5$$

$$\text{Moles in bottoms} = 162.8 - 159.5 = 3.3$$

For Ethane:

$$\text{Moles in overhead} = 2.4 \text{ (100\% to overhead)}$$

Since the isobutane (the heavy key) is 1% of the overhead stream, the sum of propane and ethane must be 99% (all n-C₄ and C₅+ are in the bottoms). Thus:

$$\text{Total Overhead Moles} = \frac{159.5 + 2.4}{0.99} = \frac{161.9}{0.99} = 163.5$$

$$\text{Moles of iC}_4 = 163.5 - 161.9 = 1.6$$

The overall balance is:

Comp.	Feed	Overhead		Bottoms	
	Moles	Moles	Mole %	Moles	Mole %
C ₂	2.4	2.4	1.5	—	—
C ₃	162.8	159.5	97.5	3.3	1.8
iC ₄	31.0	1.6	1.0	29.4	15.8
nC ₄	76.7	—	—	76.7	41.2
C ₅	76.5	—	—	76.5	41.2
Total	349.4	163.5	100.0	185.9	100.0

In actual operation, the lighter than light key components and heavier than heavy key components will not be perfectly separated. For estimation purposes and hand calculations, perfect non-key separation is a useful simplifying assumption.

Key Parameters

Two important considerations which affect the size and cost of a fractionation column are degree of separation and component volatility.

The degree of separation or product purity has a direct impact on the size of the column and the required utilities. Higher purity will require more trays, more reflux, larger diameter, and/or a reduced product quantity. One quantitative measure of the difficulty of a separation is the separation factor, S_F , defined as:

$$S_F = \left(\frac{X_D}{X_B} \right)_{LK} \left(\frac{X_B}{X_D} \right)_{HK} \quad \text{Eq 19-1}$$

Note that Equation 19-1 defines the specification for the tower design.

Typically, for most fractionation problems this factor ranges from around 500 to 2000. However, for sharp separations, it can be in the 10,000 range. The number of trays will be roughly proportional to the log of the separation factor for a given system.

The volatility of the components is usually expressed as relative volatility, α . This quantity is computed as the ratio of the equilibrium K-values of two components at a given temperature and pressure. For fractionation calculations the α of the key components is important. Therefore:

$$\alpha = K_{LK}/K_{HK} \quad \text{Eq 19-2}$$

This is a measure of the ease of separation. The larger α is, the easier is the separation.

DESIGN CONSIDERATIONS

Operating Pressure

Before any design calculations can be made on a fractionation problem, a tower operating pressure must be determined. One of the primary considerations for operating pressure is the cooling medium available for the reflux condenser. The overhead product will be at bubble point conditions for a liquid product or at dew point conditions for a vapor product. The bubble point (or dew point) pressure is fixed by a desired component separation and the temperature of the cooling medium.

The cooling media typically used are air, water, and refrigerant. Air cooling is normally the least expensive cooling method.

Practical exchanger design limits the process to a 10°C approach to the ambient summer temperature. This translates to a process temperature of 46 to 52°C in most locations. With cooling water, process temperatures of 35 to 41°C are possible. Below about 35°C, mechanical refrigeration must be used to achieve the desired condensing temperature. This is the most expensive cooling method from both a capital and operating cost standpoint. Generally, it is desirable to operate at as low a pressure as possible to maximize the relative volatility between the key components of the separation. However, if reducing the pressure requires a change to a more expensive cooling method, this is usually not a desirable choice.

In some cases, the overhead from the column must be compressed to sales or another process unit. In this case a higher operating pressure may be desired to reduce compression power.

Other items must be considered which will limit pressure selection. If an operating pressure is too high, the critical temperature of the bottom product may be exceeded and the desired separation cannot be achieved. Additionally, the pressure cannot exceed the critical pressure of the desired overhead product.

The selection of a partial or total condenser is fixed by the disposition of the overhead product. A total condenser is used for a liquid product and a partial condenser for a vapor product. However, a liquid product can be produced as a vapor and subsequently cooled and/or compressed to produce a liquid product. There may be cases where this downstream liquefaction is economically attractive. In most cases, the fractionation system for a partial condenser will be cheaper and will have to be balanced against the cost of additional downstream equipment. Before a reliable economic comparison can be made, the column design must be made for each type condenser for a number of reflux ratios and operating pressures.

Reflux Ratio and Number of Stages

The design of a fractionation column is a capital cost versus energy cost trade-off problem. The primary parameters are the number of stages and the reflux ratio. Reflux ratio may be defined in several ways. For most calculations, reflux ratio is defined as the ratio of the molar rate of reflux liquid divided by the molar rate of net overhead product. The reboiler duty is a direct function of the reflux ratio as the fractionating column must maintain an overall heat and material balance for a given separation.

A fractionation column can only produce a desired separation between the limits of minimum reflux and minimum stages. Fig. 19-6 illustrates the relationship between reflux ratio and number of stages for a given separation. At minimum reflux an infinite number of stages is required. At total reflux a minimum number of stages is required. Neither of these situations represents actual operation but are the extreme limits of possible design configurations. Methods have been developed to calculate both these cases in a rigorous manner.^{1, 2} However, these methods require a computer solution of tray-by-tray calculations. In order to begin a detailed design, estimates of minimum reflux ratio and minimum trays should be generated from simple methods using a binary key component analysis.

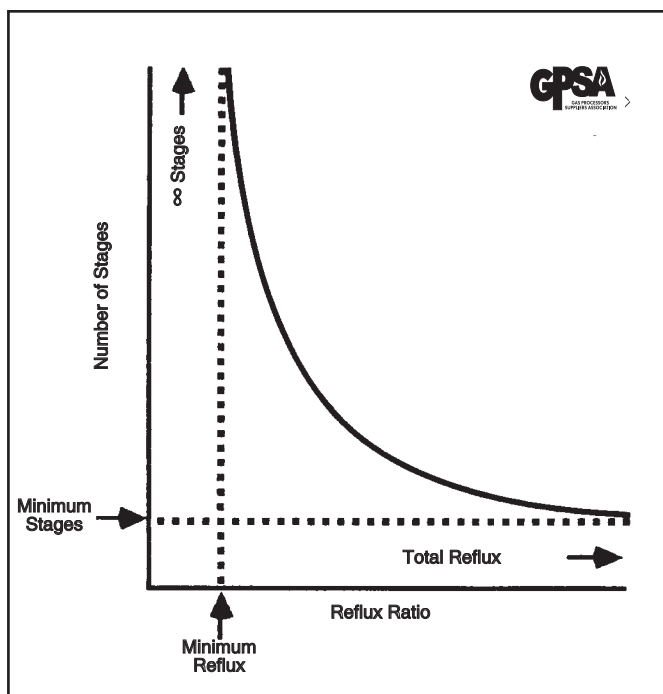
Minimum Stages

The minimum stages can be calculated for most multicomponent systems by the Fenske equation.³

$$S_m = \frac{\log S_F}{\log (\alpha_{avg})} \quad \text{Eq 19-3}$$

FIG. 19-6

Relationship Between Reflux Ratio and Number of Stages



S_m in this equation includes a partial reboiler and a partial condenser if these items are used.

The α_{avg} is the column average relative volatility of the key components in the separation. Various averaging techniques have been proposed such as square root averaging of the α at the top and bottom of the column. The most common approach is to use an arithmetic average:

$$\alpha_{avg} = \frac{\alpha_{top} + \alpha_{bottom}}{2} \quad \text{Eq 19-4}$$

If volatility varies widely, the approach of Winn⁴ is suggested, in which a modified volatility is used:

$$\beta_{ij} = K_{iLK}/K_{jHK}^b \quad \text{Eq 19-5}$$

where the exponent b is obtained from K -value plots over the range of interest.

The minimum stage calculation is:

$$S_m = \frac{\log \left[\left(\frac{X_D}{X_B} \right)_{LK} \left(\frac{X_B}{X_D} \right)_{HK}^b \left(\frac{B}{D} \right)^{1-b} \right]}{\log \beta_{ij}} \quad \text{Eq 19-6}$$

Note that S_m includes the partial condenser and partial reboiler if they exist.

Minimum Reflux Ratio

The Underwood method⁵ is the most widely used of the methods for calculating minimum reflux ratio. Underwood assumed constant relative volatility and constant molal overflow in the development of this method. The first step is to evaluate θ by trial and error:

$$1 - q = \sum_{i=1}^n \frac{X_{Fi}}{(\alpha_i - \theta)/\alpha_i} \quad \text{Eq 19-7}$$

Once θ is determined, the minimum reflux ratio is:

$$(L_0/D)_m + 1 = R_m + 1 = \sum_{i=1}^n \frac{X_{Di}}{(\alpha_i - \theta)/\alpha_i} \quad \text{Eq 19-8}$$

Number of Stages

The number of theoretical stages required for a given separation at a reflux ratio between minimum and total reflux can be determined from empirical relationships. Erbar and Maddox⁶ made an extensive investigation of tray-by-tray fractionator calculations and developed the correlation in Fig. 19-7. This correlation relates the ratio of minimum stages to theoretical stages to the minimum reflux ratio, R_m , and the operating reflux ratio, R (where $R = L_0/D$).

Fig. 19-7 can be used to determine an operating reflux for a given number of stages by entering the figure at the value of S_m/S , moving up to the line representing the value of $R_m/(R_m + 1)$ and reading a value of $R/(R + 1)$. The optimum operating reflux ratio has been found to be near the minimum reflux ratio. Values of 1.2 to 1.3 times the minimum are common.⁷ Thus for a given R , a value of S can be determined from Fig. 19-7.

This correlation is generated on the basis of bubble point feed. If the feed is between the bubble point and dew point then the operating reflux should be corrected. Erbar and Maddox⁶ proposed the following relationship to adjust the vapor rate from the top tray for non-bubble point feed:

$$V_{corr} = V_{calc} + \frac{\left(1 - \frac{D}{F}\right)[F(H_{VF} - H_{BP})]}{\left(\frac{Q_C}{I_0}\right)_{calc}} \quad \text{Eq 19-9}$$

The reflux rate can then be adjusted by material balance since:

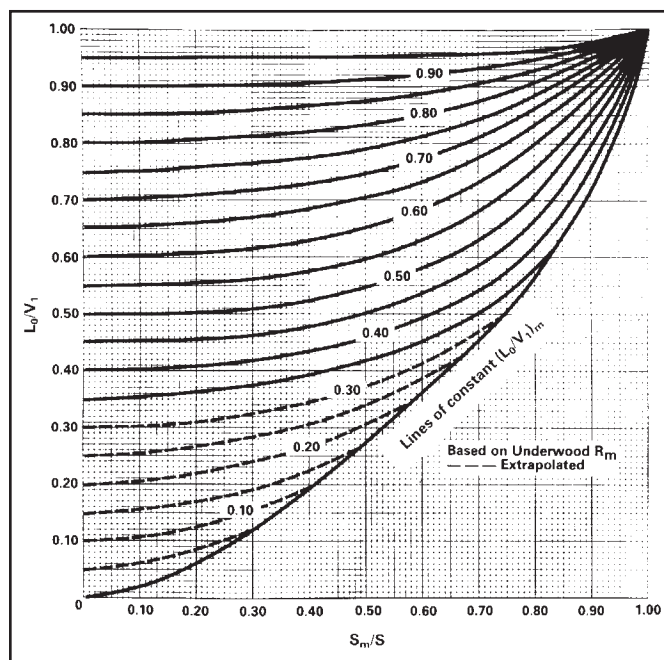
$$L_0 = V_1 - D \quad \text{Eq 19-10}$$

Computation Method

In order to determine the design parameters for a fractionation problem, the following method is recommended:

1. Establish feed composition, flow rate, temperature, and pressure.
2. Make product splits for the column and establish condenser temperature and column pressure. From column pressure, calculate the reboiler temperature.
3. Calculate minimum number of theoretical stages from the Fenske equation (Equation 19-3).
4. Calculate minimum reflux rate from the Underwood equations (Equation 19-7 and 19-8).
5. Obtain theoretical stages/operating reflux relation from Fig. 19-7.
6. Adjust actual reflux for feed vaporization if necessary (Equation 19-9, 19-10).

FIG. 19-7
Erbar-Maddox Correlation of Stages vs Reflux⁶



Example 19-2 — For the given feed stream, 1102 m³/day (bubble point feed).

Desire: 98% C₃ in the overhead (relative to the feed)
1% iC₄ in the overhead

Air cooling (49°C Condensing Temperature)

Feed Composition	Mol %	Moles/hr
C ₂	2.07	21.5
C ₃	48.67	505.6
iC ₄	10.11	105.0
nC ₄	24.08	250.1
iC ₅	5.41	56.2
nC ₅	4.81	50.0
C ₆	4.85	50.4
	100.00	1038.8

Find the:

- Minimum trays required
- Minimum reflux ratio
- Actual trays at 1.3 times the minimum reflux ratio

Solution Steps

Estimate Product Splits from Material Balance:

	Overhead		Bottoms	
	Moles	Mol %	Moles	Mol %
C ₂	21.5	4.1	—	—
C ₃	495.4	94.9	10.2	2.0
iC ₄	5.2	1.0	99.8	19.3
nC ₄	—	—	250.1	48.4
iC ₅	—	—	56.2	10.9
nC ₅	—	—	50.0	9.7
C ₆	—	—	50.4	9.7
Totals	522.1	100.0	516.7	100.0

Get tower pressure at 49°C (bubble point calculation). Using K-values from Section 25:

	K	α
C ₂	2.80	
C ₃	0.93	2.067
iC ₄	0.45	

bubble point pressure = 1930 kPa

Estimate the bottom temperature using K-values at 280 psia (bubble point calculation) assuming negligible pressure drop:

	K	α
C ₃	2.30	1.643
iC ₄	1.40	
nC ₄	1.15	
iC ₅	0.68	
nC ₅	0.62	
C ₆	0.15	

bubble point temperature = 121°C

$$\alpha_{\text{avg}} = 1.855$$

Determine the minimum number of trays (Equation 19-3):

$$S_F = \left(\frac{94.9}{2.0} \right) \left(\frac{19.3}{1.0} \right) = 915.8$$

$$S_m = \frac{\log(915.8)}{\log(1.855)} = 11 \text{ trays}$$

Correct for change in relative volatility by using Equation 19-6:

$$K_{LK} = 0.93 = \beta_{ij} (0.45)^b \text{ (condenser)}$$

$$K_{LK} = 2.30 = \beta_{ij} (1.4)^b \text{ (reboiler)}$$

dividing gives $2.473 = (3.111)^b$

$$b = 0.798; \beta_{ij} = 1.759$$

$$S_m = \frac{\log \left[\left(\frac{495.4}{10.2} \right) \left(\frac{99.8}{5.2} \right)^{0.798} \left(\frac{516.7}{522.1} \right)^{1-0.798} \right]}{\log(1.759)}$$

$$S_m = \frac{\log(512.1)}{\log(1.759)} = 11.05$$

Thus correcting for changing α did not affect the minimum tray calculation in this example.

Find the minimum reflux, R_m (Equation 19-7, 8):

	X_F	α_{avg} relative to C_6+	$\alpha x_F/(a - \theta)$		
			$\theta = 16$	$\theta = 15$	$\theta = 15.8$
C_2	0.0207	68.33	0.0270	0.0265	0.0269
C_3	0.4867	26.67	1.2165	1.1123	1.1941
iC_4	0.1011	13.83	-0.6443	-1.1951	-0.7098
nC_4	0.2408	11.00	-0.4506	-0.6622	-0.5518
iC_5	0.0541	5.83	-0.0310	-0.0344	-0.0316
nC_5	0.0481	5.00	-0.0219	-0.0241	-0.0223
C_6	0.0485	1.00	-0.0032	-0.0035	-0.0033
Total			0.0925	-0.7805	-0.0978

$$\theta = 15.9$$

$$R_m + 1 = \frac{68.33 (0.041)}{68.33 - 15.9} + \frac{26.67 (0.949)}{26.67 - 15.9} + \frac{13.83 (0.01)}{13.83 - 15.9}$$

$$R_m + 1 = 2.336$$

$$R_m = 1.336$$

Theoretical trays at $R = (1.3) R_m = 1.737$

$$L_o/V_1 = \frac{R}{R + 1} = \frac{1.737}{(1.737 + 1.0)} = 0.635$$

$$(L_o/V_1)_m = \frac{R_m}{R_m + 1} = \frac{1.336}{(1.336 + 1.0)} = 0.572$$

$$S_m/S = 0.54 \text{ (Fig. 19-7)}$$

$$S = 20.46 \text{ trays (use 21 trays)}$$

TRAYED COLUMNS

Internals

Various types of trays are used in fractionation columns. Fig. 19-8 presents an isometric sketch of the top two trays in a fractionation column with bubble caps. The bubble caps, along with the weirs and downcomers, maintain a liquid level on the trays. The liquid flows across the tray, into the downcomer, and across the next tray in the opposite direction. The vapor flows up through the caps and through the slots mixing with the liquid.

Fig. 19-9 shows the vapor flow through bubble cap trays, sieve trays, and valve trays. Due to the riser in the bubble cap, it is the only tray which can be designed to prevent liquid from “weeping” through the vapor passage. Sieve or valve trays control weeping by vapor velocity. The bubble cap tray has the highest turndown ratio, with designs of 8:1 to 10:1 ratio being common. Bubble cap trays are commonly found in existing glycol dehydration columns but new designs typically use structured packing.

Valve and sieve trays are popular due to the lower cost and increased capacity over bubble cap trays for a given tower diameter. Fig. 19-10 shows two valve designs. The upper drawing shows a floating valve free to open and close with varying vapor flow rates. The lower drawing shows a “caged” valve which prevents valve loss due to erosion of the tray. Various other designs are common such as using multiple disks and rectangular

valves. Valves of assorted weights have also been used to increase flexibility.

The sieve or perforated tray is the simplest construction of the three general types and thus is the least expensive option. The sieve tray is simply a plate with holes for vapor passage. Although the sieve tray generally has lower pressure drop than other tray types, its main disadvantage is that sieve trays will be susceptible to “weeping” or “dumping” of the liquid through the holes at low vapor rates and its turndown capacity is limited.

Vendor-proprietary fixed valves and mini-valves are also available for increased capacity. Fixed valves can provide turndown capacity between that of sieve trays and standard (movable) valves. Larger fixed valves are used as alternatives to sieve trays in fouling services. Examples include

Koch-Glitsch PROVALVE®, VG-10 (fixed valves)

VG-0 (fixed mini-valve)

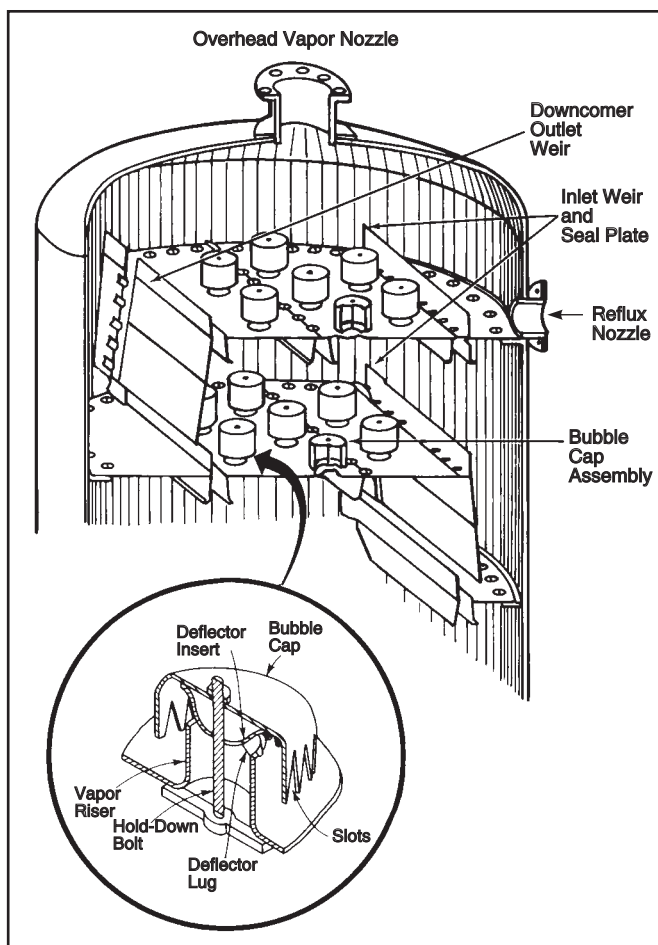
MV-1 (movable mini-valve)

Sulzer SVG (fixed valve)

MVG (fixed mini-valve)

Trayed columns generally provide satisfactory operation over a wide range of vapor and liquid loadings. Fig. 19-11 shows

FIG. 19-8
Top Two Trays of a Bubble-cap Column²⁸



operating characteristics for a representative system. The vapor and liquid rates can vary independently over a broad range and the column will operate satisfactorily. At low vapor rates unsatisfactory tray dynamics may be characterized by vapor pulsation, dumping of liquid, or uneven distribution. At high vapor rates, the tower will eventually flood as liquid is entrained to the tray above or backed-up in the downcomers. At low liquid

FIG. 19-9
Flow Through Vapor Passages²⁸

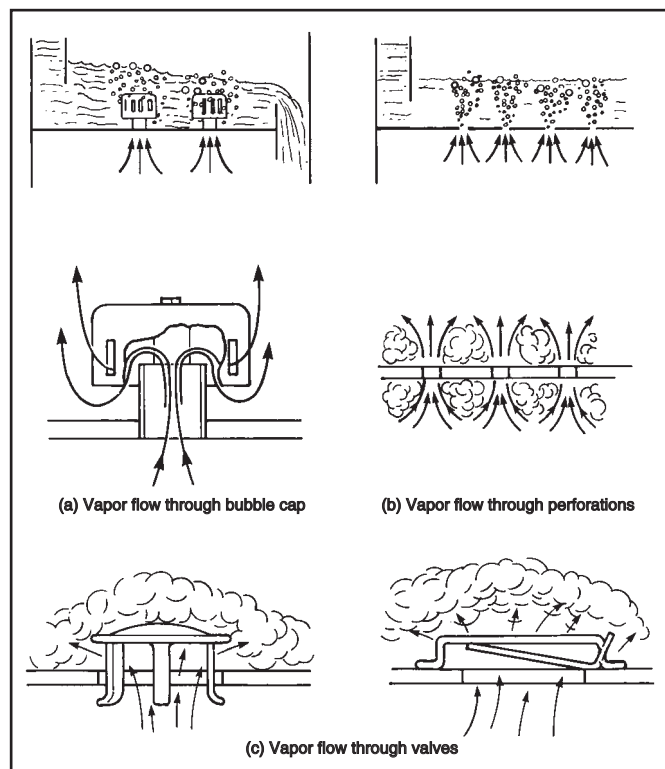
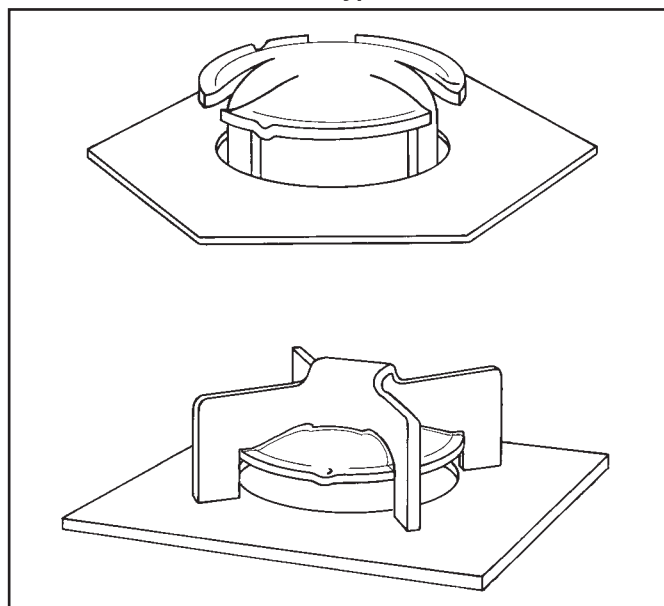


FIG. 19-10
Valve Types²⁸



rates, poor vapor-liquid contact can result. High liquid rates can cause flooding and dumping as the liquid capacity of the downcomers is exceeded.

In order to handle higher liquid rates, more downcomer area is required. This is often achieved by using multiple pass trays. Multipass trays increase liquid handling capacity for a given diameter due to increased weir length and reductions in the weir crest. Fig. 19-12 shows various configurations beyond a one pass tray where the liquid phase is split into two to four flow paths to increase liquid handling capacity.

Sizing

“C” Factor Method — Many design methods for sizing trayed fractionators have been used. Generally these methods are oriented toward liquid entrainment limitations or correlations for flooding limits. A simple method called the Souders and Brown equation⁸ involves using a Stokes’ Law type formula:

$$V_{\max} = (0.3048) C \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} \quad \text{Eq 19-11}$$

Note that ρ_L and ρ_v are at flowing temperature and pressure.

The value of C can be found from Fig. 19-13 based on tray spacing and liquid surface tension. The column diameter is:

$$D_T = \sqrt{\frac{V_{\max}}{V_{\max} (0.7854)}} \quad \text{Eq 19-12}$$

This method was originally developed for bubble cap trays and gives a conservative diameter, especially for other types of trays.

Nomograph Method — Manufacturers of valve trays have developed design methods for their trays. Design procedures are made available^{9, 10, 11} for preliminary studies. One such procedure starts with the nomograph in Fig. 19-14.¹⁰ This is a simple relationship of liquid rate (GPM) and a quantity V_{load} defined as:

$$V_{\text{load}} = \text{CFS} \sqrt{\frac{\rho_v}{(\rho_L - \rho_v)}} \quad \text{Eq 19-13}$$

FIG. 19-11

Limits of Satisfactory Tray Operation for a Specific Set of Tray Fluid Properties⁸

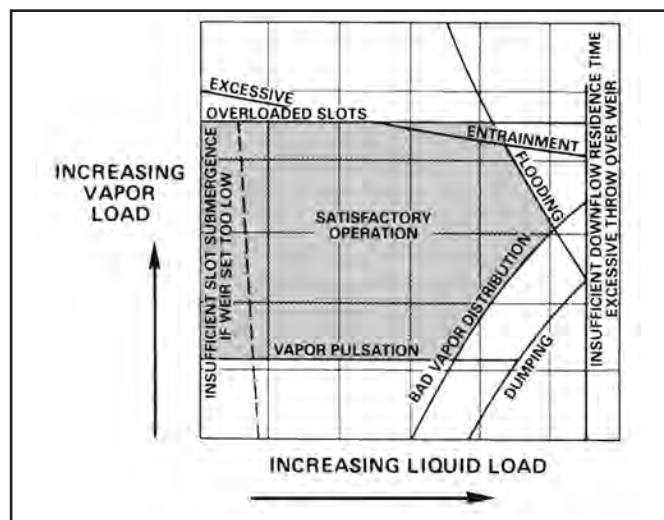
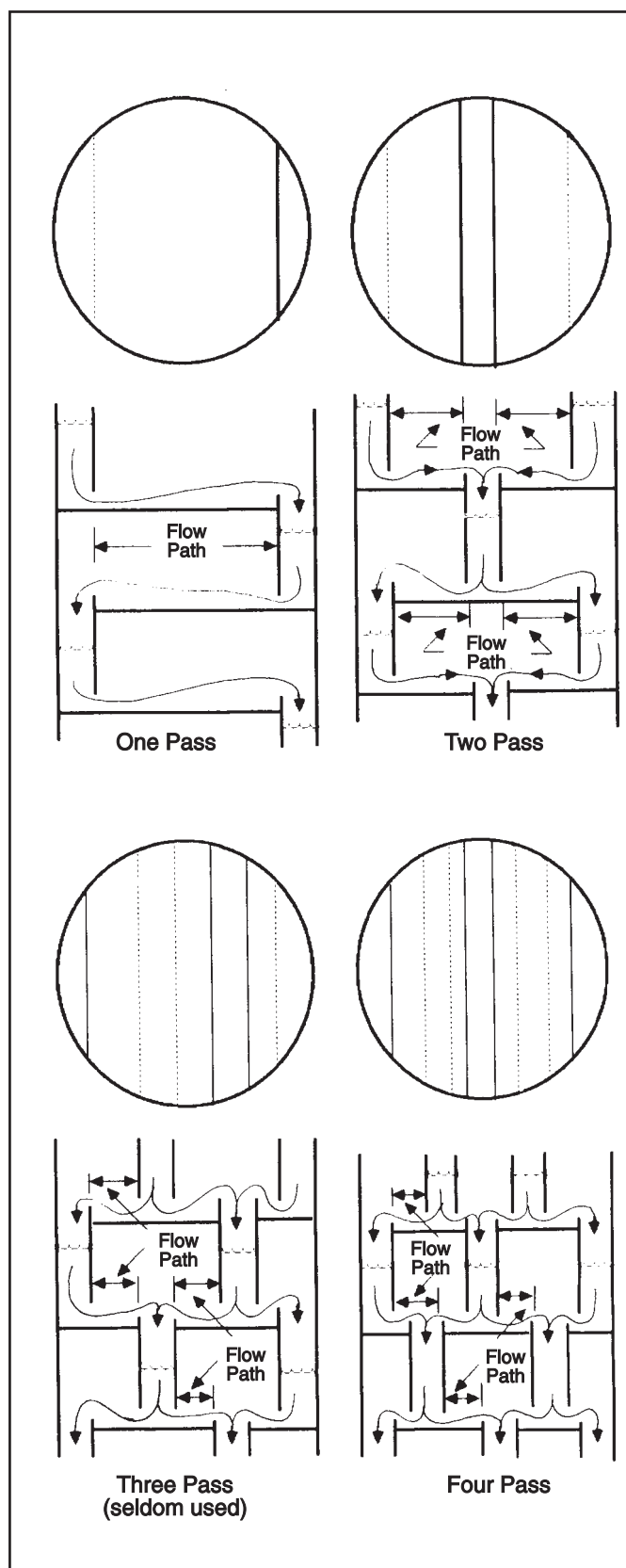


FIG. 19-12
Alternative Liquid Flow Paths



Simplified Hand Method — Tray vendors today provide computer programs to users to size both trayed and packed columns. These vendors should be contacted for copies of their programs for their products. The method included here is a hand method that can be used for preliminary sizing of trayed columns and to understand the key parameters that affect column sizing.

Fig. 19-14 is an approximation only and does not take into account foaming which is a major consideration in many systems. In order to compensate for foaming, a System Factor is used to adjust the vapor and liquid capacities (Fig. 19-15).

The downcomer velocity VD_{dsg}^* is found from Fig. 19-16. VD_{dsg}^* is corrected by the System Factor:

$$VD_{dsg} = VD_{dsg}^* (\text{System Factor}) \quad \text{Eq 19-14}$$

The other factor required for this design method is the vapor capacity factor CAF.

$$CAF = CAF_0 (\text{System Factor}) \quad \text{Eq 19-15}$$

CAF_0 is read from Fig. 19-17. In order to compute the column cross sectional area, three quantities are needed.

FIG. 19-13

Souders-Brown Correlation for Approximate Tower Sizing⁸

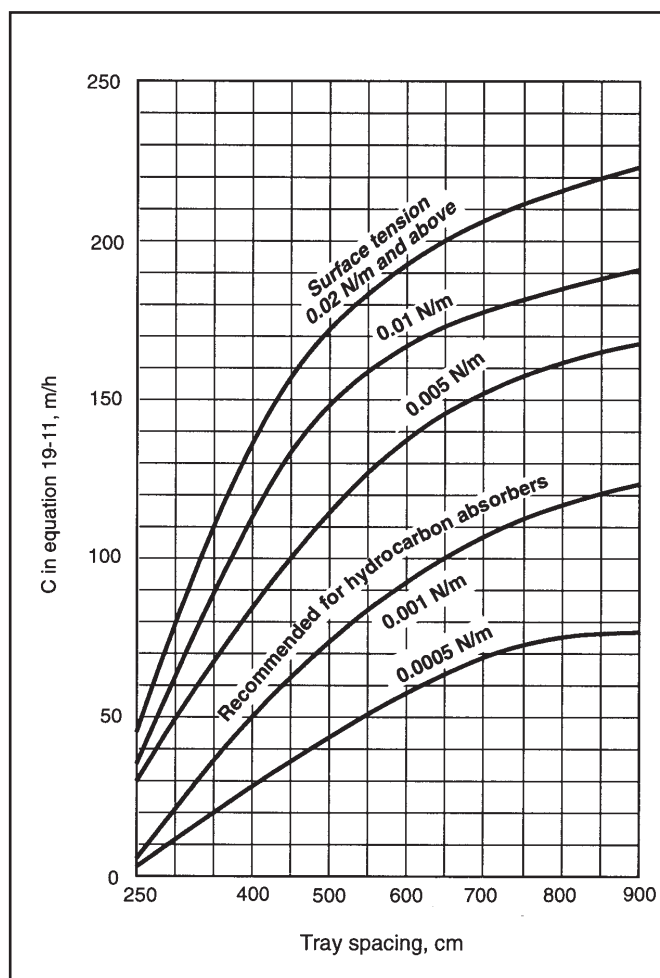


FIG. 19-14
Valve Tray Diameter¹⁰

