

Example 11.13

Calculate the plate efficiency for the plate design considered in Examples 11.11 and 11.12, using Van Winkle's correlation.

Solution

From Examples 11.12 and 11.11:

$$\rho_L = 925 \text{ kg/m}^3,$$

$$\rho_v = 1.35 \text{ kg/m}^3,$$

$$\mu_L = 0.34 \times 10^{-3} \text{ Ns/m}^2,$$

$$\mu_v = 10.0 \times 10^{-6} \text{ Ns/m}^2,$$

$$D_{LK} = D_L = 4.64 \times 10^{-9} \text{ m}^2/\text{s},$$

$$h_w = 50 \text{ mm},$$

$$FA \text{ (fractional area)} = A_h/A_c = \frac{0.038}{0.50} = 0.076,$$

$$u_v = \text{superficial vapour velocity} = \frac{0.81}{0.50} = 1.62 \text{ m/s},$$

$$\sigma_L = 60 \times 10^{-3} \text{ N/m}$$

$$Dg = \left(\frac{60 \times 10^{-3}}{0.34 \times 10^{-3} \times 1.62} \right) = 109$$

$$Sc = \left(\frac{0.34 \times 10^{-3}}{925 \times 4.64 \times 10^{-9}} \right) = 79,$$

$$Re = \left(\frac{50 \times 10^{-3} \times 1.62 \times 1.35}{0.34 \times 10^{-3} \times 0.076} \right) = 4232$$

$$E_{mV} = 0.07(109)^{0.14}(79)^{0.25}(4232)^{0.08} \quad (11.69)$$

$$= \underline{\underline{0.79}} \text{ (79 per cent)}$$

11.14. PACKED COLUMNS

Packed columns are used for distillation, gas absorption, and liquid-liquid extraction; only distillation and absorption will be considered in this section. Stripping (desorption) is the reverse of absorption and the same design methods will apply.

The gas liquid contact in a packed bed column is continuous, not stage-wise, as in a plate column. The liquid flows down the column over the packing surface and the gas or vapour, counter-currently, up the column. In some gas-absorption columns co-current flow is used. The performance of a packed column is very dependent on the maintenance of good liquid and gas distribution throughout the packed bed, and this is an important consideration in packed-column design.

A schematic diagram, showing the main features of a packed absorption column, is given in Figure 11.36. A packed distillation column will be similar to the plate columns shown in Figure 11.1, with the plates replaced by packed sections.

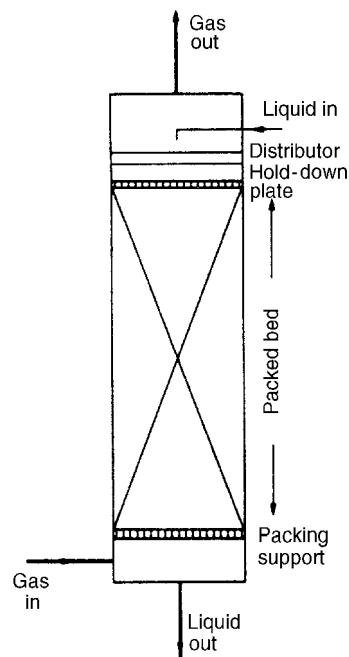


Figure 11.36. Packed absorption column

The design of packed columns using random packings is covered in books by Strigle (1994) and Billet (1995).

Choice of plates or packing

The choice between a plate or packed column for a particular application can only be made with complete assurance by costing each design. However, this will not always be worthwhile, or necessary, and the choice can usually be made, on the basis of experience by considering main advantages and disadvantages of each type; which are listed below:

1. Plate columns can be designed to handle a wider range of liquid and gas flow-rates than packed columns.
2. Packed columns are not suitable for very low liquid rates.
3. The efficiency of a plate can be predicted with more certainty than the equivalent term for packing (HETP or HTU).
4. Plate columns can be designed with more assurance than packed columns. There is always some doubt that good liquid distribution can be maintained throughout a packed column under all operating conditions, particularly in large columns.

5. It is easier to make provision for cooling in a plate column; coils can be installed on the plates.
6. It is easier to make provision for the withdrawal of side-streams from plate columns.
7. If the liquid causes fouling, or contains solids, it is easier to make provision for cleaning in a plate column; manways can be installed on the plates. With small-diameter columns it may be cheaper to use packing and replace the packing when it becomes fouled.
8. For corrosive liquids a packed column will usually be cheaper than the equivalent plate column.
9. The liquid hold-up is appreciably lower in a packed column than a plate column. This can be important when the inventory of toxic or flammable liquids needs to be kept as small as possible for safety reasons.
10. Packed columns are more suitable for handling foaming systems.
11. The pressure drop per equilibrium stage (HETP) can be lower for packing than plates; and packing should be considered for vacuum columns.
12. Packing should always be considered for small diameter columns, say less than 0.6 m, where plates would be difficult to install, and expensive.

Packed-column design procedures

The design of a packed column will involve the following steps:

1. Select the type and size of packing.
2. Determine the column height required for the specified separation.
3. Determine the column diameter (capacity), to handle the liquid and vapour flow rates.
4. Select and design the column internal features: packing support, liquid distributor, redistributors.

These steps are discussed in the following sections, and a packed-column design illustrated in Example 11.14.

11.14.1. Types of packing

The principal requirements of a packing are that it should:

- Provide a large surface area: a high interfacial area between the gas and liquid.
- Have an open structure: low resistance to gas flow.
- Promote uniform liquid distribution on the packing surface.
- Promote uniform vapour gas flow across the column cross-section.

Many diverse types and shapes of packing have been developed to satisfy these requirements. They can be divided into two broad classes:

1. Packings with a regular geometry: such as stacked rings, grids and proprietary structured packings.
2. Random packings: rings, saddles and proprietary shapes, which are dumped into the column and take up a random arrangement.

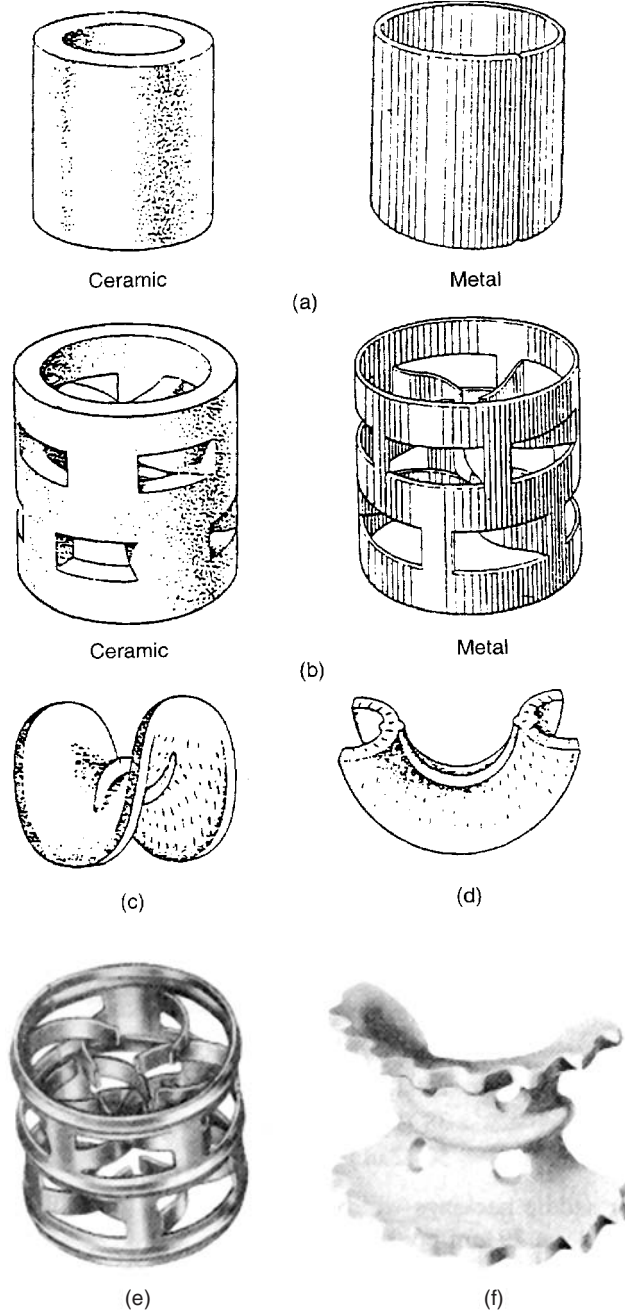


Figure 11.37. Types of packing (Norton Co.). (a) Raschig rings (b) Pall rings (c) Berl saddle ceramic (d) Intalox saddle ceramic (e) Metal Hypac (f) Ceramic, super Intalox

Grids have an open structure and are used for high gas rates, where low pressure drop is essential; for example, in cooling towers. Random packings and structured packing elements are more commonly used in the process industries.

Random packing

The principal types of random packings are shown in Figure 11.37 (see p. 590). Design data for these packings are given in Table 11.2. Data on a wider range of packing sizes are given in Volume 2, Chapter 4. The design methods and data given in this section can be used for the preliminary design of packed columns, but for detailed design it is advisable to consult the packing manufacturer's technical literature to obtain data for the particular packing that will be used. The packing manufacturers should be consulted for details of the many special types of packing that are available for special applications.

Raschig rings, Figure 11.37*a*, are one of the oldest specially manufactured types of random packing, and are still in general use. Pall rings, Figure 11.37*b*, are essentially Raschig rings in which openings have been made by folding strips of the surface into the ring. This increases the free area and improves the liquid distribution characteristics. Berl saddles, Figure 11.37*c*, were developed to give improved liquid distribution compared to Raschig rings, Intalox saddles, Figure 11.37*d*, can be considered to be an improved type of Berl saddle; their shape makes them easier to manufacture than Berl saddles. The Hypac and Super Intalox packings shown in Figure 11.37*e*, *f* can be considered improved types of Pall ring and Intalox saddle, respectively.

Table 11.2. Design data for various packings

	Size		Bulk density (kg/m ³)	Surface area a (m ² /m ³)	Packing factor F_p m ⁻¹
	in.	mm			
Raschig rings	0.50	13	881	368	2100
ceramic	1.0	25	673	190	525
	1.5	38	689	128	310
	2.0	51	651	95	210
	3.0	76	561	69	120
Metal	0.5	13	1201	417	980
(density for carbon steel)	1.0	25	625	207	375
	1.5	38	785	141	270
	2.0	51	593	102	190
	3.0	76	400	72	105
Pall rings	0.625	16	593	341	230
metal	1.0	25	481	210	160
(density for carbon steel)	1.25	32	385	128	92
	2.0	51	353	102	66
	3.5	76	273	66	52
Plastics	0.625	16	112	341	320
(density for polypropylene)	1.0	25	88	207	170
	1.5	38	76	128	130
	2.0	51	68	102	82
	3.5	89	64	85	52
Intalox saddles	0.5	13	737	480	660
ceramic	1.0	25	673	253	300
	1.5	38	625	194	170
	2.0	51	609	108	130
	3.0	76	577		72

Intalox saddles, Super Intalox and Hypac packings are proprietary design, and registered trade marks of the Norton Chemical Process Products Ltd.

Ring and saddle packings are available in a variety of materials: ceramics, metals, plastics and carbon. Metal and plastics (polypropylene) rings are more efficient than ceramic rings, as it is possible to make the walls thinner.

Raschig rings are cheaper per unit volume than Pall rings or saddles but are less efficient, and the total cost of the column will usually be higher if Raschig rings are specified. For new columns, the choice will normally be between Pall rings and Berl or Intalox saddles.

The choice of material will depend on the nature of the fluids and the operating temperature. Ceramic packing will be the first choice for corrosive liquids; but ceramics are unsuitable for use with strong alkalis. Plastics packings are attacked by some organic solvents, and can only be used up to moderate temperatures; so are unsuitable for distillation columns. Where the column operation is likely to be unstable metal rings should be specified, as ceramic packing is easily broken. The choice of packings for distillation and absorption is discussed in detail by Eckert (1963), Strigle (1994), Kister (1992) and Billet (1995).

Packing size

In general, the largest size of packing that is suitable for the size of column should be used, up to 50 mm. Small sizes are appreciably more expensive than the larger sizes. Above 50 mm the lower cost per cubic metre does not normally compensate for the lower mass transfer efficiency. Use of too large a size in a small column can cause poor liquid distribution.

Recommended size ranges are:

Column diameter	Use packing size
<0.3 m (1 ft)	<25 mm (1 in.)
0.3 to 0.9 m (1 to 3 ft)	25 to 38 mm (1 to 1.5 in.)
>0.9 m	50 to 75 mm (2 to 3 in.)

Structured packing

The term *structured packing* refers to packing elements made up from wire mesh or perforated metal sheets. The material is folded and arranged with a regular geometry, to give a high surface area with a high void fraction. A typical example is shown in Figure 11.38.

Structured packings are produced by a number of manufacturers. The basic construction and performance of the various proprietary types available are similar. They are available in metal, plastics and stoneware. The advantage of structured packings over random packing is their low HETP (typically less than 0.5 m) and low pressure drop (around 100 Pa/m). They are being increasingly used in the following applications:

1. For difficult separations, requiring many stages: such as the separation of isotopes.
2. High vacuum distillation.
3. For column revamps: to increase capacity and reduce reflux ratio requirements.

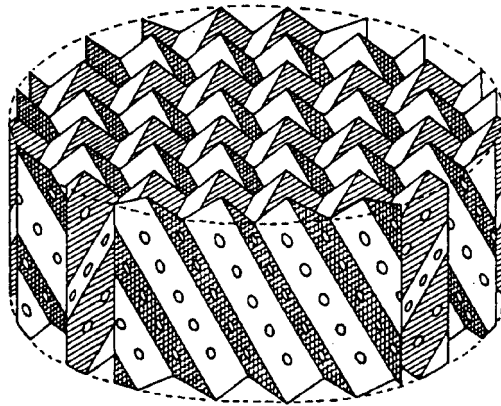


Figure 11.38. Make-up of structured packing. (Reproduced from Butcher (1988) with permission.)

The applications have mainly been in distillation, but structured packings can also be used in absorption; in applications where high efficiency and low pressure drop are needed.

The cost of structured packings per cubic metre will be significantly higher than that of random packings, but this is offset by their higher efficiency.

The manufacturers' technical literature should be consulted for design data. A review of the types available is given by Butcher (1988). Generalised methods for predicting the capacity and pressure drop of structured packings are given by Fair and Bravo (1990) and Kister and Gill (1992). The use of structured packings in distillation is discussed in detail in the book by Kister (1992).

11.14.2. Packed-bed height

Distillation

For the design of packed distillation columns it is simpler to treat the separation as a staged process, and use the concept of the height of an equivalent equilibrium stage to convert the number of ideal stages required to a height of packing. The methods for estimating the number of ideal stages given in Sections 11.5 to 11.8 can then be applied to packed columns.

The height of an equivalent equilibrium stage, usually called the height of a theoretical plate (HETP), is the height of packing that will give the same separation as an equilibrium stage. It has been shown by Eckert (1975) that in distillation the HETP for a given type and size of packing is essentially constant, and independent of the system physical properties; providing good liquid distribution is maintained and the pressure drop is at least above 17 mm water per metre of packing height. The following values for Pall rings can be used to make an approximate estimate of the bed height required.

Size, mm	HETP, m
25 (1 in.)	0.4–0.5
38 ($1\frac{1}{2}$ in.)	0.6–0.75
50 (2 in.)	0.75–1.0

The HETP for saddle packings will be similar to that for Pall rings providing the pressure drop is at least 29 mm per m.

The HETP for Raschig rings will be higher than those for Pall rings or saddles, and the values given above will only apply at an appreciably higher pressure drop, greater than 42 mm per m.

The methods for estimating the heights of transfer units, HTU, given in Section 11.14.3 can be used for distillation. The relationship between transfer units and the height of an equivalent theoretical plate, HETP is given by:

$$\text{HETP} = \frac{\mathbf{H}_{OG} \text{Ln} \left(\frac{mG_m}{L_m} \right)}{\left(\frac{mG_m}{L_m - 1} \right)} \quad (11.96)$$

from equation 11.105

$$\mathbf{H}_{OG} = \mathbf{H}_G + \left(\frac{mG_m}{L_m} \right) \mathbf{H}_L$$

The slope of the operating line m will normally vary throughout a distillation so it will be necessary to calculate the HETP for each plate or a series of plates.

Absorption

Though packed absorption and stripping columns can also be designed as staged process, it is usually more convenient to use the integrated form of the differential equations set up by considering the rates of mass transfer at a point in the column. The derivation of these equations is given in Volume 2, Chapter 12.

Where the concentration of the solute is small, say less than 10 per cent, the flow of gas and liquid will be essentially constant throughout the column, and the height of packing required, Z , is given by:

$$Z = \frac{G_m}{K_G a P} \int_{y_2}^{y_1} \frac{dy}{y - y_e} \quad (11.97)$$

in terms of the overall gas phase mass transfer coefficient K_G and the gas composition. Or,

$$Z = \frac{L_m}{K_L a C_t} \int_{x_2}^{x_1} \frac{dx}{x_e - x} \quad (11.98)$$

in terms of the overall liquid-phase mass-transfer coefficient K_L and the liquid composition,

where G_m = molar gas flow-rate per unit cross-sectional area,

L_m = molar liquid flow-rate per unit cross-sectional area,

a = interfacial surface area per unit volume,

P = total pressure,

C_t = total molar concentration,

y_1 and y_2 = the mol fractions of the solute in the gas at the bottom and top of the column, respectively,

x_1 and x_2 = the mol fractions of the solute in the liquid at the bottom and top of the column, respectively,

x_e = the concentration in the liquid that would be in equilibrium with the gas concentration at any point,

y_e = the concentration in the gas that would be in equilibrium with the liquid concentration at any point.

The relation between the equilibrium concentrations and actual concentrations is shown in Figure 11.39.

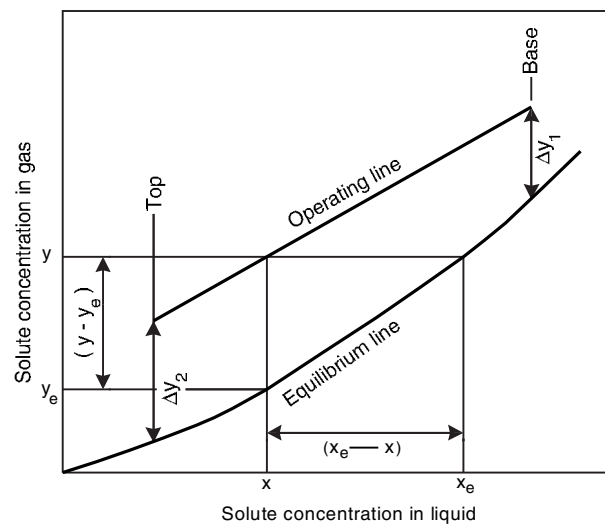


Figure 11.39. Gas absorption concentration relationships

For design purposes it is convenient to write equations 11.97 and 11.98 in terms of “transfer units” (HTU); where the value of integral is the number of transfer units, and the group in front of the integral sign, which has units of length, is the height of a transfer unit.

$$Z = \mathbf{H}_{OG} \mathbf{N}_{OG} \quad (11.99a)$$

or

$$Z = \mathbf{H}_{OL} \mathbf{N}_{OL} \quad (11.99b)$$

where \mathbf{H}_{OG} is the height of an overall gas-phase transfer unit

$$= \frac{G_m}{K_G a P} \quad (11.100)$$

\mathbf{N}_{OG} is the number of overall gas-phase transfer units

$$= \int_{y_2}^{y_1} \frac{dy}{y - y_e} \quad (11.101)$$

\mathbf{H}_{OL} is the height of an overall liquid-phase transfer unit

$$= \frac{L_m}{K_L a C_t} \quad (11.102)$$

N_{OL} is the number of overall liquid-phase transfer units

$$= \int_{x_2}^{x_1} \frac{dx}{x_e - x} \quad (11.103)$$

The number of overall gas-phase transfer units is often more conveniently expressed in terms of the partial pressure of the solute gas.

$$N_{OG} = \int_{p_1}^{p_2} \frac{dp}{p - p_e} \quad (11.104)$$

The relationship between the overall height of a transfer unit and the individual film transfer units H_L and H_G , which are based on the concentration driving force across the liquid and gas films, is given by:

$$H_{OG} = H_G + m \frac{G_m}{L_m} H_L \quad (11.105)$$

$$H_{OL} = H_L + \frac{L_m}{mG_m} H_G \quad (11.106)$$

where m is the slope of the equilibrium line and G_m/L_m the slope of the operating line.

The number of transfer units is obtained by graphical or numerical integration of equations 11.101, 11.103 or 11.104.

Where the operating and equilibrium lines are straight, and they can usually be considered to be so for dilute systems, the number of transfer units is given by:

$$N_{OG} = \frac{y_1 - y_2}{\Delta y_{lm}} \quad (11.107)$$

where Δy_{lm} is the log mean driving force, given by:

$$y_{lm} = \frac{\Delta y_1 - \Delta y_2}{\ln \left(\frac{\Delta y_1}{\Delta y_2} \right)} \quad (11.108)$$

where $\Delta y_1 = y_1 - y_e$,

$\Delta y_2 = y_2 - y_e$.

If the equilibrium curve and operating lines can be taken as straight and the solvent feed essentially solute free, the number of transfer units is given by:

$$N_{OG} = \frac{1}{1 - \left(\frac{mG_m}{L_m} \right)} \ln \left[\left(1 - \frac{mG_m}{L_m} \right) \frac{y_1}{y_2} + \frac{mG_m}{L_m} \right] \quad (11.109)$$

This equation is plotted in Figure 11.40, which can be used to make a quick estimate of the number of transfer units required for a given separation.

It can be seen from Figure 11.40 that the number of stages required for a given separation is very dependent on the flow rate L_m . If the solvent rate is not set by

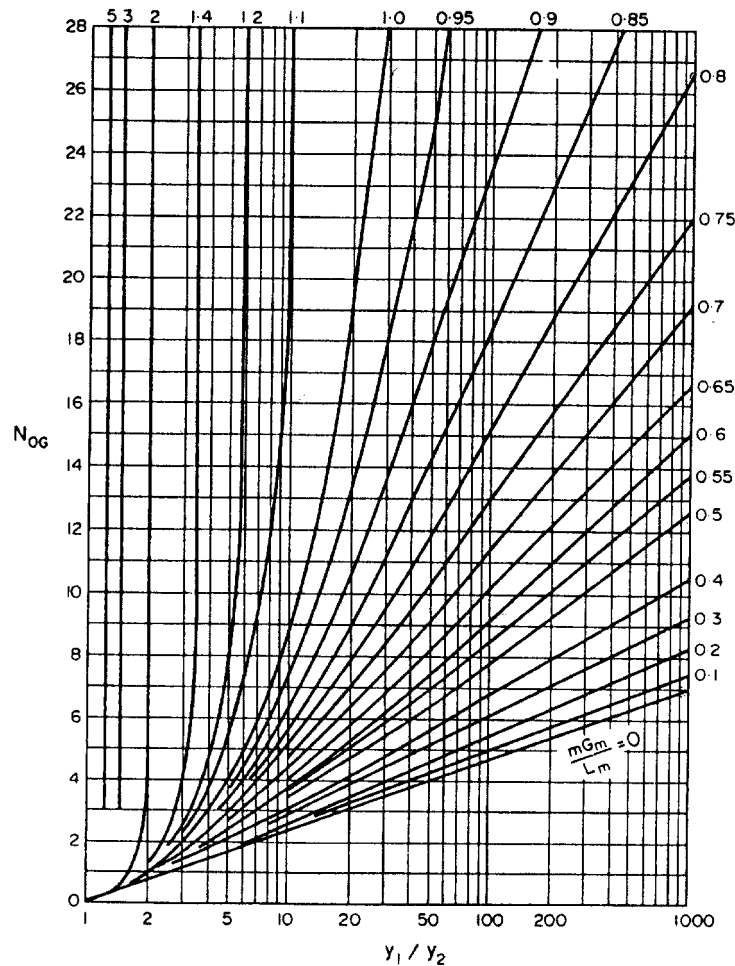


Figure 11.40. Number of transfer units N_{OG} as a function of y_1/y_2 with mG_m/L_m as parameter

other process considerations, Figure 11.40 can be used to make quick estimates of the column height at different flow rates to find the most economic value. Colburn (1939) has suggested that the optimum value for the term mG_m/L_m will lie between 0.7 to 0.8.

Only physical absorption from dilute gases has been considered in this section. For a discussion of absorption from concentrated gases and absorption with chemical reaction, the reader should refer to Volume 2, or to the book by Treybal (1980). If the inlet gas concentration is not too high, the equations for dilute systems can be used by dividing the operating line up into two or three straight sections.

11.14.3. Prediction of the height of a transfer unit (HTU)

There is no entirely satisfactory method for predicting the height of a transfer unit. In practice the value for a particular packing will depend not only on the physical properties

and flow-rates of the gas and liquid, but also on the uniformity of the liquid distribution throughout the column, which is dependent on the column height and diameter. This makes it difficult to extrapolate data obtained from small size laboratory and pilot plant columns to industrial size columns. Whenever possible estimates should be based on actual values obtained from operating columns of similar size to that being designed.

Experimental values for several systems are given by Cornell *et al.* (1960), Eckert (1963), and Vital *et al.* (1984). A selection of values for a range of systems is given in Table 11.3. The composite mass transfer term $K_G a$ is normally used when reporting experimental mass-transfer coefficients for packing, as the effective interfacial area for mass transfer will be less than the actual surface area a of the packing.

Many correlations have been published for predicting the height of a transfer unit, and the mass-transfer coefficients; several are reviewed in Volume 2, Chapter 12. The two methods given in this section have been found to be reliable for preliminary design work, and, in the absence of practical values, can be used for the final design with a suitable factor of safety.

The approach taken by the authors of the two methods is fundamentally different, and this provides a useful cross-check on the predicted values. Judgement must always be used when using predictive methods in design, and it is always worthwhile trying several methods and comparing the results.

Typical values for the HTU of random packings are:

25 mm (1 in.)	0.3 to 0.6 m (1 to 2 ft)
38 mm ($1\frac{1}{2}$ in.)	0.5 to 0.75 m ($1\frac{1}{2}$ to $2\frac{1}{2}$ ft)
50 mm (2 in.)	0.6 to 1.0 m (2 to 3 ft)

Table 11.3. Typical packing efficiencies

System	Pressure kPa	Column dia, m	Packing type	Packing size, mm	HTU m	HETP m
<i>Absorption</i>						
Hydrocarbons	6000	0.9	Pall	50		0.85
NH ₃ -Air-H ₂ O	101	—	Berl	50	0.50	
Air-water	101	—	Berl	50	0.50	
Acetone-water	101	0.6	Pall	50		0.75
<i>Distillation</i>						
Pentane-propane	101	0.46	Pall	25		0.46
IPA-water	101	0.46	Int.	25	0.75	0.50
Methanol-water	101	0.41	Pall	25	0.52	
	101	0.20	Int.	25		0.46
Acetone-water	101	0.46	Pall	25		0.37
	101	0.36	Int.	25		0.46
Formic acid-water	101	0.91	Pall	50		0.45
Acetone-water	101	0.38	Pall	38	0.55	0.45
	101	0.38	Int.	50	0.50	0.45
	101	1.07	Int.	38		1.22
MEK-toluene	101	0.38	Pall	25	0.29	0.35
	101	0.38	Int.	25	0.27	0.23
	101	0.38	Berl	25	0.31	0.31

Pall = Pall rings, Berl = Berl saddles, Int. = Intalox saddles

Cornell's method

Cornell *et al.* (1960) reviewed the previously published data and presented empirical equations for predicting the height of the gas and liquid film transfer units. Their correlation takes into account the physical properties of the system, the gas and liquid flow-rates; and the column diameter and height. Equations and figures are given for a range of sizes of Raschig rings and Berl saddles. Only those for Berl saddles are given here, as it is unlikely that Raschig rings would be considered for a new column. Though the mass-transfer efficiency of Pall rings and Intalox saddles will be higher than that of the equivalent size Berl saddle, the method can be used to make conservative estimates for these packings.

Bolles and Fair (1982) have extended the correlations given in the earlier paper to include metal Pall rings.

Cornell's equations are:

$$\mathbf{H}_G = 0.011 \psi_h (Sc)_v^{0.5} \left(\frac{D_c}{0.305} \right)^{1.11} \left(\frac{Z}{3.05} \right)^{0.33} / (L_w^* f_1 f_2 f_3)^{0.5} \quad (11.110)$$

$$\mathbf{H}_L = 0.305 \phi_h (Sc)_L^{0.5} K_3 \left(\frac{Z}{3.05} \right)^{0.15} \quad (11.111)$$

where \mathbf{H}_G = height of a gas-phase transfer unit, m,

\mathbf{H}_L = height of a liquid-phase transfer unit, m,

$(Sc)_v$ = gas Schmidt number = $(\mu_v / \rho_v D_v)$,

$(Sc)_L$ = liquid Schmidt number = $(\mu_L / \rho_L D_L)$,

D_c = column diameter, m,

Z = column height, m,

K_3 = percentage flooding correction factor, from Figure 11.41,

ψ_h = \mathbf{H}_G factor from Figure 11.42,

ϕ_h = \mathbf{H}_L factor from Figure 11.43,

L_w^* = liquid mass flow-rate per unit area column cross-sectional area, $\text{kg/m}^2\text{s}$,

f_1 = liquid viscosity correction factor = $(\mu_L / \mu_w)^{0.16}$,

f_2 = liquid density correction factor = $(\rho_w / \rho_L)^{1.25}$,

f_3 = surface tension correction factor = $(\sigma_w / \sigma_L)^{0.8}$,

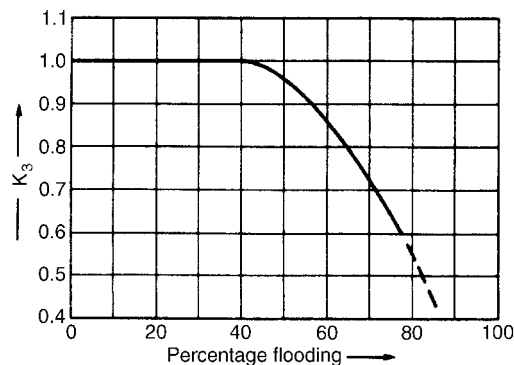


Figure 11.41. Percentage flooding correction factor

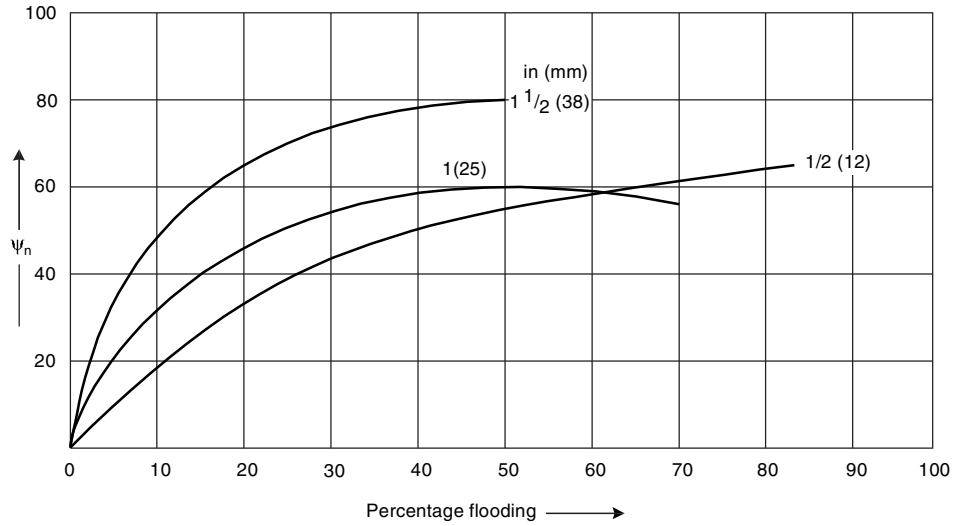


Figure 11.42. Factor for H_G for Berl saddles

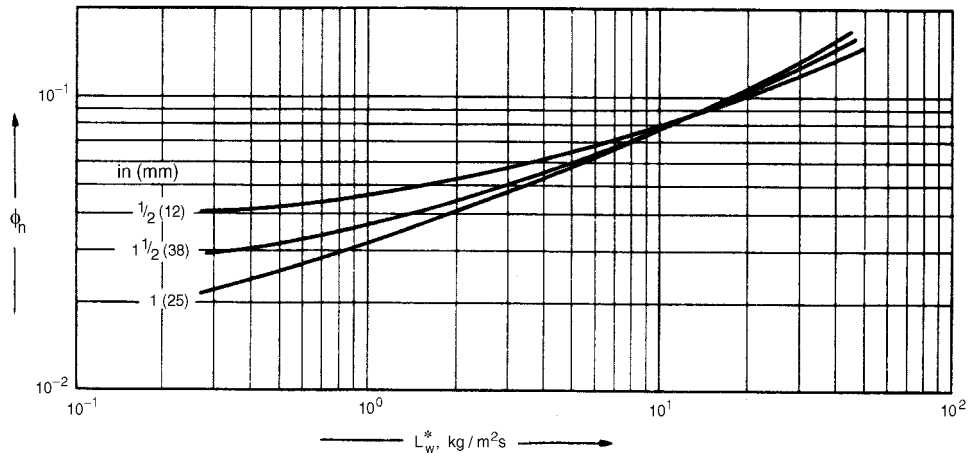


Figure 11.43. Factor for H_L for Berl saddles

where the suffix w refers to the physical properties of water at 20°C; all other physical properties are evaluated at the column conditions.

The terms $(D_c/0.305)$ and $(Z/3.05)$ are included in the equations to allow for the effects of column diameter and packed-bed height. The “standard” values used by Cornell were 1 ft (0.305 m) for diameter, and 10 ft (3.05 m) for height. These correction terms will clearly give silly results if applied over too wide a range of values. For design purposes the diameter correction term should be taken as a fixed value of 2.3 for columns above 0.6 m

(2 ft) diameter, and the height correction should only be included when the distance between liquid redistributors is greater than 3 m. To use Figures 11.41 and 11.42 an estimate of the column percentage flooding is needed. This can be obtained from Figure 11.44, where a flooding line has been included with the lines of constant pressure drop.

$$\text{Percentage flooding} = \left[\frac{K_4 \text{ at design pressure drop}}{K_4 \text{ at flooding}} \right]^{1/2} \quad (11.112)$$

A full discussion of flooding in packed columns is given in Volume 2, Chapter 4.

Onda's method

Onda *et al.* (1968) published useful correlations for the film mass-transfer coefficients k_G and k_L and the effective wetted area of the packing a_w , which can be used to calculate H_G and H_L .

Their correlations were based on a large amount of data on gas absorption and distillation; with a variety of packings, which included Pall rings and Berl saddles. Their method for estimating the effective area of packing can also be used with experimentally determined values of the mass-transfer coefficients, and values predicted using other correlations.

The equation for the effective area is:

$$\frac{a_w}{a} = 1 - \exp \left[-1.45 \left(\frac{\sigma_c}{\sigma_L} \right)^{0.75} \left(\frac{L_w^*}{a\mu_L} \right)^{0.1} \left(\frac{L_w^{*2}a}{\rho_L^2 g} \right)^{-0.05} \left(\frac{L_w^{*2}}{\rho_L \sigma_L a} \right)^{0.2} \right] \quad (11.113)$$

and for the mass coefficients:

$$k_L \left(\frac{\rho_L}{\mu_L g} \right)^{1/3} = 0.0051 \left(\frac{L_w^*}{a_w \mu_L} \right)^{2/3} \left(\frac{\mu_L}{\rho_L D_L} \right)^{-1/2} (ad_p)^{0.4} \quad (11.114)$$

$$\frac{k_G RT}{a D_v} = K_5 \left(\frac{V_w^*}{a\mu_v} \right)^{0.7} \left(\frac{\mu_v}{\rho_v D_v} \right)^{1/3} (ad_p)^{-2.0} \quad (11.115)$$

where $K_5 = 5.23$ for packing sizes above 15 mm, and 2.00 for sizes below 15 mm,

L_w^* = liquid mass flow rate per unit cross-sectional area, kg/m²s,

V_w^* = gas mass flow rate per unit column cross-sectional area, kg/m²s,

a_w = effective interfacial area of packing per unit volume, m²/m³,

a = actual area of packing per unit volume (see Table 11.3), m²/m³,

d_p = packing size, m,

σ_c = critical surface tension for the particular packing material given below:

Material	σ_c mN/m
Ceramic	61
Metal (steel)	75
Plastic (polyethylene)	33
Carbon	56

σ_L = liquid surface tension, N/m,

k_G = gas film mass transfer coefficient, kmol/m²s atm or kmol/m²s bar,

k_L = liquid film mass transfer coefficient, kmol/m²s (kmol/m³) = m/s.

Note: all the groups in the equations are dimensionless.

The units for k_G will depend on the units used for the gas constant:

$$\mathbf{R} = 0.08206 \text{ atm m}^3/\text{kmol K or}$$

$$0.08314 \text{ bar m}^3/\text{kmol K}$$

The film transfer unit heights are given by:

$$\mathbf{H}_G = \frac{G_m}{k_G a_w P} \quad (11.116)$$

$$\mathbf{H}_L = \frac{L_m}{k_L a_w C_t} \quad (11.117)$$

where P = column operating pressure, atm or bar,

C_t = total concentration, kmol/m³ = ρ_L /molecular weight solvent,

G_m = molar gas flow-rate per unit cross-sectional area, kmol/m²s,

L_m = molar liquid flow-rate per unit cross-sectional area, kmol/m²s.

Nomographs

A set of nomographs are given in Volume 2, Chapter 12 for the estimation of \mathbf{H}_G and \mathbf{H}_L , and the wetting rate. These were taken from a proprietary publication, but are based on a set of similar nomographs given by Czermann *et al.* (1958), who developed the nomographs from correlations put forward by Morris and Jackson (1953) and other workers.

The nomographs can be used to make a quick, rough, estimate of the column height, but are an oversimplification, as they do not take into account all the physical properties and other factors that affect mass transfer in packed columns.

11.14.4. Column diameter (capacity)

The capacity of a packed column is determined by its cross-sectional area. Normally, the column will be designed to operate at the highest economical pressure drop, to ensure good liquid and gas distribution. For random packings the pressure drop will not normally exceed 80 mm of water per metre of packing height. At this value the gas velocity will be about 80 per cent of the flooding velocity. Recommended design values, mm water per m packing, are:

Absorbers and strippers	15 to 50
Distillation, atmospheric and moderate pressure	40 to 80

Where the liquid is likely to foam, these values should be halved.

For vacuum distillations the maximum allowable pressure drop will be determined by the process requirements, but for satisfactory liquid distribution the pressure drop should not be less than 8 mm water per m. If very low bottom pressures are required special low pressure-drop gauze packings should be considered; such as Hyperfil, Multifil or Dixon rings; see Volume 2, Chapter 4.

The column cross-sectional area and diameter for the selected pressure drop can be determined from the generalised pressure-drop correlation given in Figure 11.44.

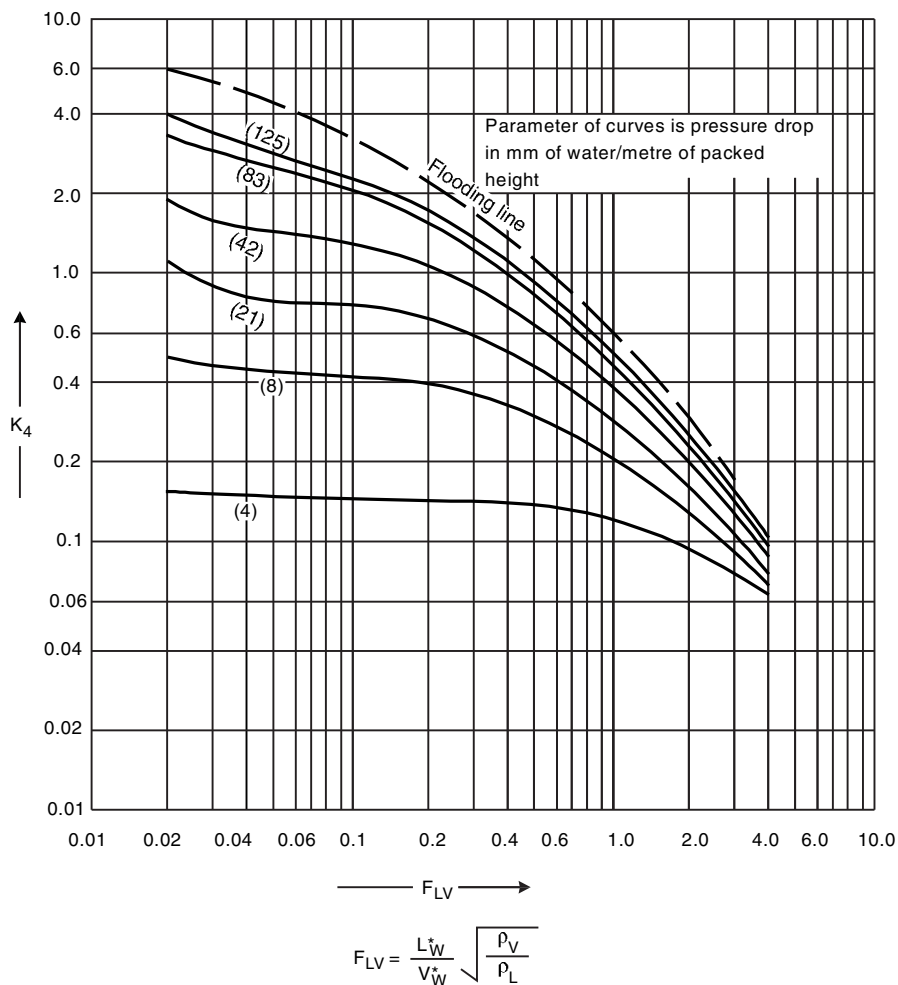


Figure 11.44. Generalised pressure drop correlation, adapted from a figure by the Norton Co. with permission

Figure 11.44 correlates the liquid and vapour flow rates, system physical properties and packing characteristics, with the gas mass flow-rate per unit cross-sectional area; with lines of constant pressure drop as a parameter.

The term K_4 on Figure 11.44 is the function:

$$K_4 = \frac{13.1(V_w^*)^2 F_p \left(\frac{\mu_L}{\rho_L}\right)^{0.1}}{\rho_v(\rho_L - \rho_v)} \quad (11.118)$$

where V_w^* = gas mass flow-rate per unit column cross-sectional area, kg/m²s

F_p = packing factor, characteristic of the size and type of packing,
see Table 11.3, m⁻¹.

μ_L = liquid viscosity, Ns/m²

ρ_L, ρ_v = liquid and vapour densities, kg/m³

The values of the flow factor F_{LV} given in Figure 11.44 covers the range that will generally give satisfactory column performance.

The ratio of liquid to gas flow will be fixed by the reflux ratio in distillation; and in gas absorption will be selected to give the required separation with the most economic use of solvent.

A new generalised correlation for pressure drop in packed columns, similar to Figure 11.44, has been published by Leva (1992), (1995). The new correlations gives a better prediction for systems where the density of the irrigating fluid is appreciably greater than that of water. It can also be used to predict the pressure drop over dry packing.

Example 11.14

Sulphur dioxide produced by the combustion of sulphur in air is absorbed in water. Pure SO₂ is then recovered from the solution by steam stripping. Make a preliminary design for the absorption column. The feed will be 5000 kg/h of gas containing 8 per cent v/v SO₂. The gas will be cooled to 20°C. A 95 per cent recovery of the sulphur dioxide is required.

Solution

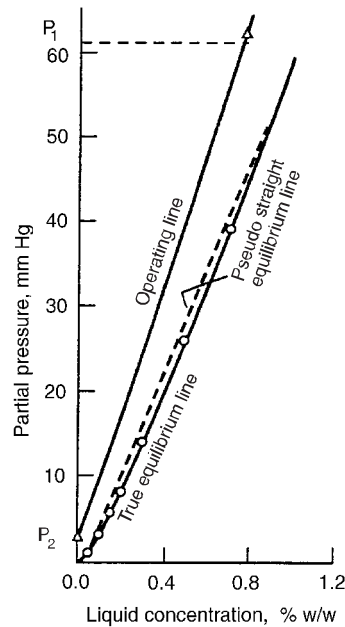
As the solubility of SO₂ in water is high, operation at atmospheric pressure should be satisfactory. The feed-water temperature will be taken as 20°C, a reasonable design value.

Solubility data

From *Chemical Engineers Handbook*, 5th edn, McGraw-Hill, 1973.

per cent w/w solution	0.05	0.1	0.15	0.2	0.3	0.5	0.7	1.0	1.5
SO ₂ Partial press. gas mmHg	1.2	3.2	5.8	8.5	14.1	26	39	59	92

Partial pressure of SO₂ in the feed = (8/100) × 760 = 60.8 mm Hg

Figure (d). SO₂ absorber design (Example 11.14)

These figures are plotted in Figure (d).

Number of stages

Partial pressure in the exit gas at 95 per cent recovery = $60.8 \times 0.05 = 3.04$ mm Hg

Over this range of partial pressure the equilibrium line is essentially straight so Figure 11.40 can be used to estimate the number of stages needed.

The use of Figure 11.40 will slightly overestimate the number of stages and a more accurate estimate would be made by graphical integration of equation 11.104; but this is not justified in view of the uncertainty in the prediction of the transfer unit height.

Molecular weights: SO₂ = 64, H₂O = 18, air = 29

Slope of equilibrium line

From the data: partial pressure at 1.0% w/w SO₂ = 59 mm Hg.

$$\text{Mol. fraction in vapour} = \frac{59}{760} = 0.0776$$

$$\text{Mol. fraction in liquid} = \frac{1}{\frac{64}{1} + \frac{99}{18}} = 0.0028$$

$$m = \frac{0.0776}{0.0028} = 27.4$$

To decide the most economic water flow-rate, the stripper design should be considered together with the absorption design, but for the purpose of this example the absorption design will be considered alone. Using Figure 11.40 the number of stages required at different water rates will be determined and the “optimum” rate chosen:

$$\frac{y_1}{y_2} = \frac{p_1}{p_2} = \frac{60.8}{3.04} = 20$$

$m \frac{G_m}{L_m}$	0.5	0.6	0.7	0.8	0.9	1.0
N_{OG}	3.7	4.1	6.3	8	10.8	19.0

It can be seen that the “optimum” will be between $mG_m/L_m = 0.6$ to 0.8 , as would be expected. Below 0.6 there is only a small decrease in the number of stages required with increasing liquid rate; and above 0.8 the number of stages increases rapidly with decreasing liquid rate.

Check the liquid outlet composition at 0.6 and 0.8 :

$$\text{Material balance } L_m x_1 = G_m (y_1 - y_2)$$

$$\text{so } x_1 = \frac{G_m}{L_m} (0.08 \times 0.95) = \frac{m}{27.4} \frac{G_m}{L_m} (0.076)$$

$$\text{at } \frac{mG_m}{L_m} = 0.6, x_1 = 1.66 \times 10^{-3} \text{ mol fraction}$$

$$\text{at } \frac{mG_m}{L_m} = 0.8, x_1 = 2.22 \times 10^{-3} \text{ mol fraction}$$

Use 0.8 , as the higher concentration will favour the stripper design and operation, without significantly increasing the number of stages needed in the absorber.

$$N_{OG} = \underline{\underline{8}}$$

Column diameter

The physical properties of the gas can be taken as those for air, as the concentration of SO_2 is low.

$$\text{Gas flow-rate} = \frac{5000}{3600} = 1.39 \text{ kg/s}, = \frac{1.39}{29} = 0.048 \text{ kmol/s}$$

$$\begin{aligned} \text{Liquid flow-rate} &= \frac{27.4}{0.8} \times 0.048 = 1.64 \text{ kmol/s} \\ &= 29.5 \text{ kg/s.} \end{aligned}$$

Select 38 mm ($1\frac{1}{2}$ in.) ceramic Intalox saddles.

From Table 11.3, $F_p = 170 \text{ m}^{-1}$

$$\text{Gas density at } 20^\circ\text{C} = \frac{29}{22.4} \times \frac{273}{293} = 1.21 \text{ kg/m}^3$$

$$\text{Liquid density} \simeq 1000 \text{ kg/m}^3$$

Liquid viscosity = 10^{-3} Ns/m²

$$\frac{L_W^*}{V_W^*} \sqrt{\frac{\rho_v}{\rho_L}} = \frac{29.5}{1.39} \sqrt{\frac{1.21}{10^3}} = 0.74$$

Design for a pressure drop of 20 mm H₂O/m packing

From Figure 11.44,

$$K_4 = 0.35$$

At flooding $K_4 = 0.8$

$$\text{Percentage flooding} = \sqrt{\frac{0.35}{0.8}} \times 100 = 66 \text{ per cent, satisfactory.}$$

From equation 11.118

$$\begin{aligned} V_W^* &= \left[\frac{K_4 \rho_V (\rho_L - \rho_v)}{13.1 F_p (\mu_L / \rho_L)^{0.1}} \right]^{1/2} \\ &= \left[\frac{0.35 \times 1.21 (1000 - 1.21)}{13.1 \times 170 (10^{-3} / 10^3)^{0.1}} \right]^{1/2} = 0.87 \text{ kg/m}^2\text{s} \end{aligned}$$

$$\text{Column area required} = \frac{1.39}{0.87} = 1.6 \text{ m}^2$$

$$\text{Diameter} = \sqrt{\frac{4}{\pi}} \times 1.6 = 1.43 \text{ m}$$

Round off to 1.50 m

$$\text{Column area} = \frac{\pi}{4} \times 1.5^2 = 1.77 \text{ m}^2$$

$$\text{Packing size to column diameter ratio} = \frac{1.5}{38 \times 10^{-3}} = 39,$$

A larger packing size could be considered.

Percentage flooding at selected diameter

$$= 66 \times \frac{1.6}{1.77} = 60 \text{ per cent,}$$

Could consider reducing column diameter.

Estimation of H_{OG}

Cornell's method

$$D_L = 1.7 \times 10^{-9} \text{ m}^2/\text{s}$$

$$D_v = 1.45 \times 10^{-5} \text{ m}^2/\text{s}$$

$$\mu_v = 0.018 \times 10^{-3} \text{ Ns/m}^2$$

$$(Sc)_v = \frac{0.018 \times 10^{-3}}{1.21 \times 1.45 \times 10^{-5}} = 1.04$$

$$(Sc)_L = \frac{10^{-3}}{1000 \times 1.7 \times 10^{-9}} = 588$$

$$L_W^* = \frac{29.5}{1.77} = 16.7 \text{ kg/s m}^2$$

From Figure 11.41, at 60 per cent flooding, $K_3 = 0.85$.

From Figure 11.42, at 60 per cent flooding, $\psi_h = 80$.

From Figure 11.43, at $L_W^* = 16.7$, $\phi_h = 0.1$.

H_{OG} can be expected to be around 1 m, so as a first estimate Z can be taken as 8 m. The column diameter is greater than 0.6 m so the diameter correction term will be taken as 2.3.

$$H_L = 0.305 \times 0.1(588)^{0.5} \times 0.85 \left(\frac{8}{3.05} \right)^{0.15} = 0.7 \text{ m} \quad (11.111)$$

As the liquid temperature has been taken as 20°C, and the liquid is water,

$$f_1 = f_2 = f_3 = 1$$

$$H_G = 0.011 \times 80(1.04)^{0.5}(2.3) \left(\frac{8}{3.05} \right)^{0.33} / (16.7)^{0.5} = 0.7 \text{ m} \quad (11.110)$$

$$H_{OG} = 0.7 + 0.8 \times 0.7 = 1.3 \text{ m} \quad (11.105)$$

$$Z = 8 \times 1.3 = 10.4 \text{ m, close enough to the estimated value.}$$

Onda's method

$$R = 0.08314 \text{ bar m}^3/\text{kmol K.}$$

Surface tension of liquid, taken as water at 20°C = 70×10^{-3} N/m

$$g = 9.81 \text{ m/s}^2$$

$$d_p = 38 \times 10^{-3} \text{ m}$$

From Table 11.3, for 38 mm Intalox saddles

$$a = 194 \text{ m}^2/\text{m}^3$$

$$\sigma_c \text{ for ceramics} = 61 \times 10^{-3} \text{ N/m}$$

$$\frac{a_W}{a} = 1 - \exp \left[-1.45 \left(\frac{61 \times 10^{-3}}{70 \times 10^{-3}} \right)^{0.75} \left(\frac{17.6}{194 \times 10^{-3}} \right)^{0.1} \left(\frac{17.6^2 \times 194}{1000^2 \times 9.81} \right)^{-0.05} \right. \\ \left. \times \left(\frac{17.6^2}{1000 \times 70 \times 10^{-3} \times 194} \right)^{0.2} \right] = 0.71 \quad (11.113)$$

$$a_W = 0.71 \times 194 = 138 \text{ m}^2/\text{m}^3$$

$$k_L \left(\frac{10^3}{10^{-3} \times 9.81} \right)^{1/3} = 0.0051 \left(\frac{17.6}{138 \times 10^{-3}} \right)^{2/3} \left(\frac{10^{-3}}{10^3 \times 1.7 \times 10^{-9}} \right)^{-1/2} \\ \times (194 \times 38 \times 10^{-3})^{0.4} \quad (11.114)$$

$$k_L = 2.5 \times 10^{-4} \text{ m/s}$$

$$V_W^* \text{ on actual column diameter} = \frac{1.39}{1.77} = 0.79 \text{ kg/m}^2\text{s}$$

$$k_G \frac{0.08314 \times 293}{194 \times 1.45 \times 10^{-5}} = 5.23 \left(\frac{0.79}{194 \times 0.018 \times 10^{-3}} \right)^{0.7} \quad (11.115)$$

$$\times \left(\frac{0.018 \times 10^{-3}}{1.21 \times 1.45 \times 10^{-5}} \right)^{1/3} (194 \times 38 \times 10^{-3})^{-2.0}$$

$$k_G = 5.0 \times 10^{-4} \text{ kmol/sm}^2 \text{ bar}$$

$$G_m = \frac{0.79}{29} = 0.027 \text{ kmol/m}^2\text{s}$$

$$L_m = \frac{16.7}{18} = 0.93 \text{ kmol/m}^2\text{s}$$

$$\mathbf{H}_G = \frac{0.027}{5.0 \times 10^{-4} \times 138 \times 1.013} = 0.39 \text{ m} \quad (11.116)$$

$$C_T = \text{total concentration, as water,}$$

$$= \frac{1000}{18} = 55.6 \text{ kmol/m}^3$$

$$\mathbf{H}_L = \frac{0.93}{2.5 \times 10^{-4} \times 138 \times 55.6} = 0.49 \text{ m} \quad (11.117)$$

$$\mathbf{H}_{OG} = 0.39 + 0.8 \times 0.49 = \underline{\underline{0.78 \text{ m}}} \quad (11.105)$$

Use higher value, estimated using Cornell's method, and round up packed bed height to 11 m.

11.14.5. Column internals

The internal fittings in a packed column are simpler than those in a plate column but must be carefully designed to ensure good performance. As a general rule, the standard fittings developed by the packing manufacturers should be specified. Some typical designs are shown in Figures 11.45 to 11.54; and their use is discussed in the following paragraphs.

Packing support

The function of the support plate is to carry the weight of the wet packing, whilst allowing free passage of the gas and liquid. These requirements conflict; a poorly designed support will give a high pressure drop and can cause local flooding. Simple grid and perforated plate supports are used, but in these designs the liquid and gas have to vie for the same openings. Wide-spaced grids are used to increase the flow area; with layers of larger size packing stacked on the grid to support the small size random packing, Figure 11.45.

The best design of packing support is one in which gas inlets are provided above the level where the liquid flows from the bed; such as the gas-injection type shown in Figure 11.46

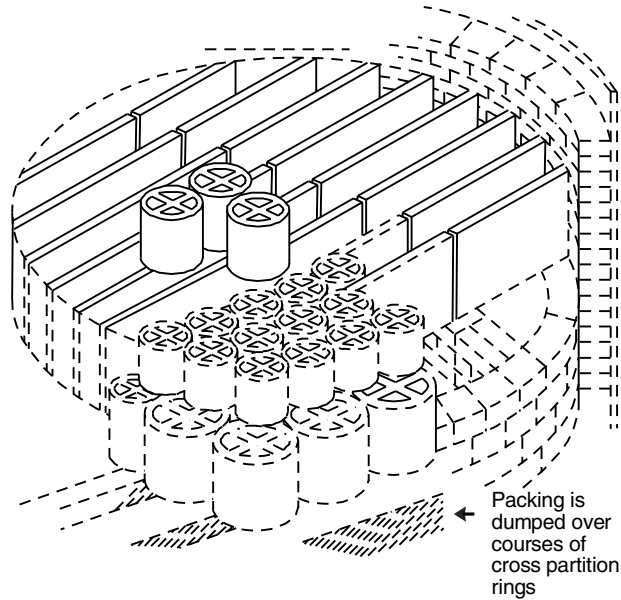


Figure 11.45. Stacked packing used to support random packing

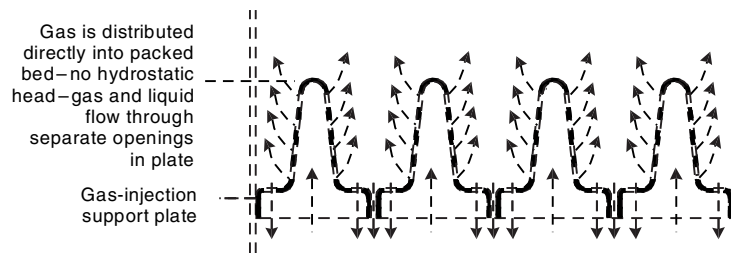


Figure 11.46. The principle of the gas-injection packing support

and 11.47. These designs have a low pressure drop and no tendency to flooding. They are available in a wide range of sizes and materials: metals, ceramics and plastics.

Liquid distributors

The satisfactory performance of a plate column is dependent on maintaining a uniform flow of liquid throughout the column, and good initial liquid distribution is essential. Various designs of distributors are used. For small-diameter columns a central open feed-pipe, or one fitted with a spray nozzle, may well be adequate; but for larger columns more elaborate designs are needed to ensure good distribution at all liquid flow-rates. The two most commonly used designs are the orifice type, shown in Figure 11.48, and the weir type, shown in Figure 11.49. In the orifice type the liquid flows through holes in the plate and the gas through short stand pipes. The gas pipes should be sized to give sufficient area for gas flow without creating a significant pressure drop; the holes should be small

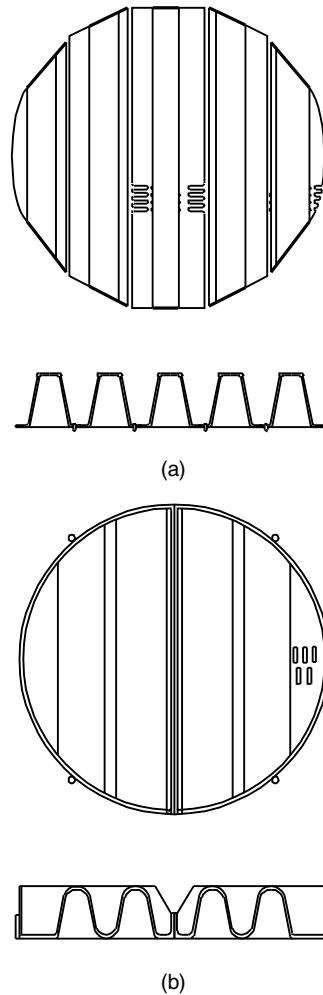


Figure 11.47. Typical designs of gas-injection supports (Norton Co.). (a) Small diameter columns (b) Large diameter columns

enough to ensure that there is a level of liquid on the plate at the lowest liquid rate, but large enough to prevent the distributor overflowing at the highest rate. In the weir type the liquid flows over notched weirs in the gas stand-pipes. This type can be designed to cope with a wider range of liquid flow rates than the simpler orifice type.

For large-diameter columns, the trough-type distributor shown in Figure 11.50 can be used, and will give good liquid distribution with a large free area for gas flow.

All distributors which rely on the gravity flow of liquid must be installed in the column level, or maldistribution of liquid will occur.

A pipe manifold distributor, Figure 11.51, can be used when the liquid is fed to the column under pressure and the flow-rate is reasonably constant. The distribution pipes and orifices should be sized to give an even flow from each element.

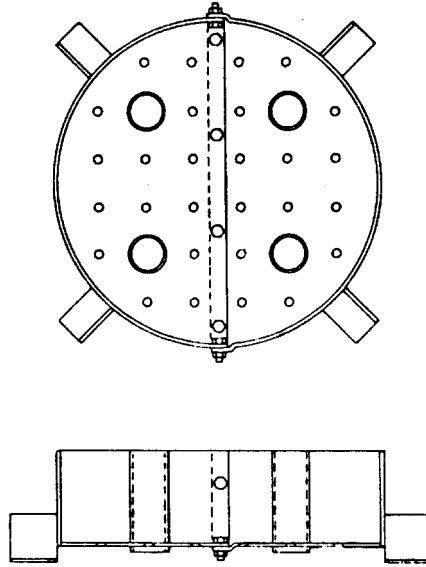


Figure 11.48. Orifice-type distributor (Norton Co.)

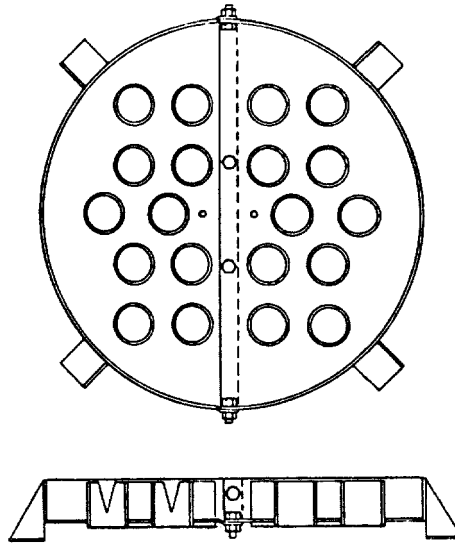


Figure 11.49. Weir-type distributor (Norton Co.)

Liquid redistributors

Redistributors are used to collect liquid that has migrated to the column walls and redistribute it evenly over the packing. They will also even out any maldistribution that has occurred within the packing.

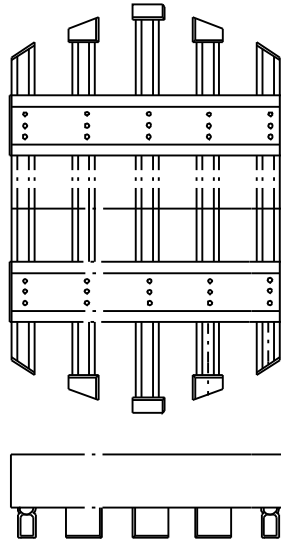


Figure 11.50. Weir-trough distributors (Norton Co.)

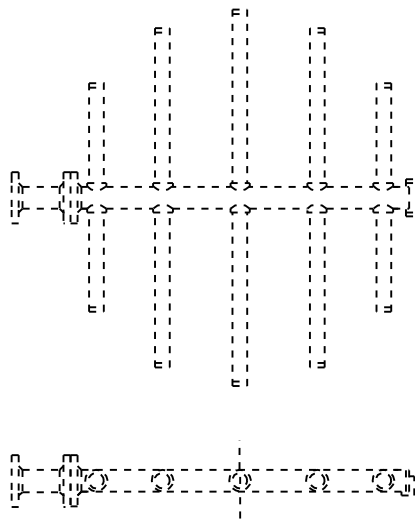


Figure 11.51. Pipe distributor (Norton Co.)

A full redistributor combines the functions of a packing support and a liquid distributor; a typical design is shown in Figure 11.52.

The “wall-wiper” type of redistributor, in which a ring collects liquid from the column wall and redirects it into the centre packing, is occasionally used in small-diameter columns, less than 0.6 m. Care should be taken when specifying this type to select a design that does not unduly restrict the gas flow and cause local flooding. A good design is that shown in Figure 11.53.

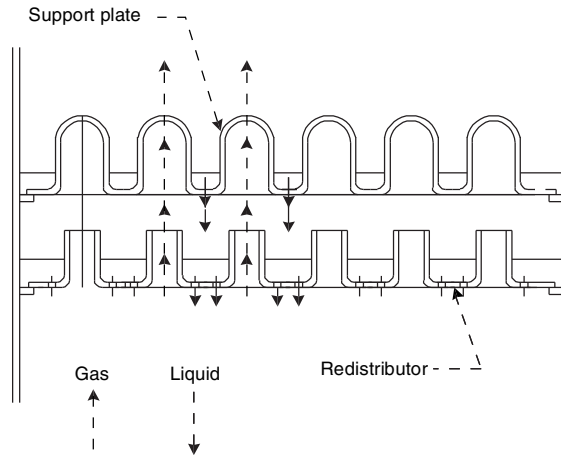


Figure 11.52. Full redistributor

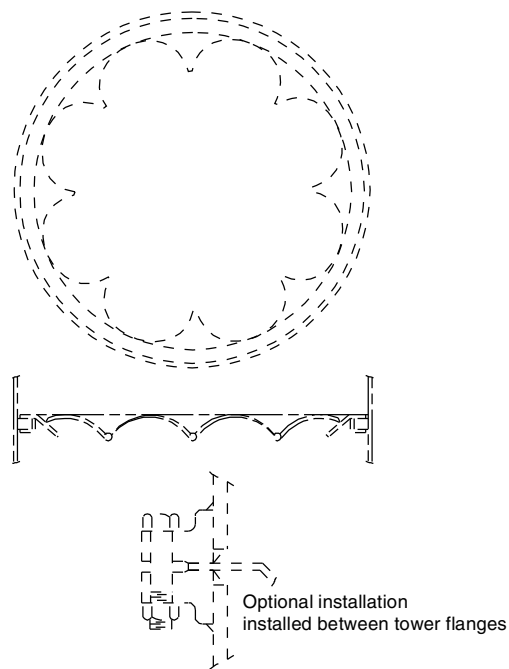


Figure 11.53. "Wall wiper" redistributor (Norton Co.)

The maximum bed height that should be used without liquid redistribution depends on the type of packing and the process. Distillation is less susceptible to maldistribution than absorption and stripping. As a general guide, the maximum bed height should not exceed 3 column diameters for Raschig rings, and 8 to 10 for Pall rings and saddles. In a large-diameter column the bed height will also be limited by the maximum weight of packing that can be supported by the packing support and column walls; this will be around 8 m.

Hold-down plates

At high gas rates, or if surging occurs through mis-operation, the top layers of packing can be fluidised. Under these conditions ceramic packing can break up and the pieces filter down the column and plug the packing; metal and plastic packing can be blown out of the column. Hold-down plates are used with ceramic packing to weigh down the top layers and prevent fluidisation; a typical design is shown in Figure 11.54. Bed-limiters are sometimes used with plastics and metal packings to prevent expansion of the bed when operating at a high-pressure drop. They are similar to hold-down plates but are of lighter construction and are fixed to the column walls. The openings in hold-down plates and bed-limiters should be small enough to retain the packing, but should not restrict the gas and liquid flow.

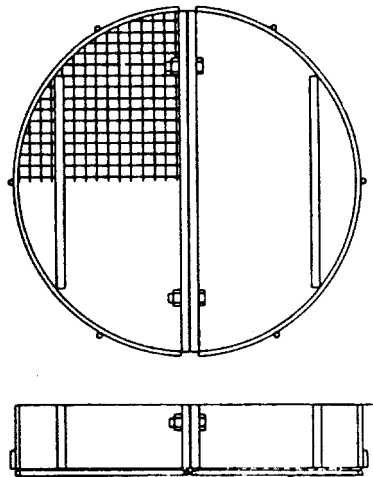


Figure 11.54. Hold-down plate design (Norton Co.)

Installing packing

Ceramic and metal packings are normally dumped into the column “wet”, to ensure a truly random distribution and prevent damage to the packing. The column is partially filled with water and the packing dumped into the water. A height of water must be kept above the packing at all times.

If the columns must be packed dry, for instance to avoid contamination of process fluids with water, the packing can be lowered into the column in buckets or other containers. Ceramic packings should not be dropped from a height of more than half a metre.

Liquid hold-up

An estimate of the amount of liquid held up in the packing under operating conditions is needed to calculate the total load carried by the packing support. The liquid hold-up will depend on the liquid rate and, to some extent, on the gas flow-rate. The packing manufacturers’ design literature should be consulted to obtain accurate estimates. As a

rough guide, a value of about 25 per cent of the packing weight can be taken for ceramic packings.

11.14.6. Wetting rates

If very low liquid rates have to be used, outside the range of F_{LV} given in Figure 11.44, the packing wetting rate should be checked to make sure it is above the minimum recommended by the packing manufacturer.

Wetting rate is defined as:

$$\text{wetting rate} = \frac{\text{volumetric liquid rate per unit cross-sectional area}}{\text{packing surface area per unit volume}}$$

A nomograph for the calculation of wetting rates is given in Volume 2, Chapter 4.

Wetting rates are frequently expressed in terms of mass or volume flow-rate per unit column cross-sectional area.

Kister (1992) gives values for minimum wetting rates of 0.5 to 2 gpm/ft² (0.35×10^3 to 1.4×10^3 m³ s⁻¹/m²) for random packing and 0.1 to 0.2 gpm/ft² (0.07×10^{-3} to 0.14×10^{-3} m³ s⁻¹/m²) for structured packing. Norman (1961) recommends that the liquid rate in absorbers should be kept above 2.7 kg/m²s.

If the design liquor rate is too low, the diameter of the column should be reduced. For some processes liquid can be recycled to increase the flow over the packing.

A substantial factor of safety should be applied to the calculated bed height for process where the wetting rate is likely to be low.

11.15. COLUMN AUXILIARIES

Intermediate storage tanks will normally be needed to smooth out fluctuations in column operation and process upsets. These tanks should be sized to give sufficient hold-up time for smooth operation and control. The hold-up time required will depend on the nature of the process and on how critical the operation is; some typical values for distillation processes are given below:

<i>Operation</i>	<i>Time, minutes</i>
Feed to a train of columns	10 to 20
Between columns	5 to 10
Feed to a column from storage	2 to 5
Reflux drum	5 to 15

The time given is that for the level in the tank to fall from the normal operating level to the minimum operating level if the feed ceases.

Horizontal or vertical tanks are used, depending on the size and duty. Where only a small hold-up volume is required this can be provided by extending the column base, or, for reflux accumulators, by extending the bottom header of the condenser.

The specification and sizing of surge tanks and accumulators is discussed in more detail by Mehra (1979) and Evans (1980).

DISTILLATION COLUMN DESIGN

As mentioned, distillation columns are designed using [VLE data](#) for the mixtures to be separated. The vapour-liquid equilibrium characteristics (indicated by the shape of the equilibrium curve) of the mixture will determine the number of stages, and hence the number of trays, required for the separation. This is illustrated clearly by applying the [McCabe-Thiele](#) method to design a binary column.

McCABE-THIELE DESIGN METHOD

The McCabe-Thiele approach is a graphical one, and uses the VLE plot to determine the theoretical number of stages required to effect the separation of a binary mixture. It assumes **constant molar overflow** and this implies that:

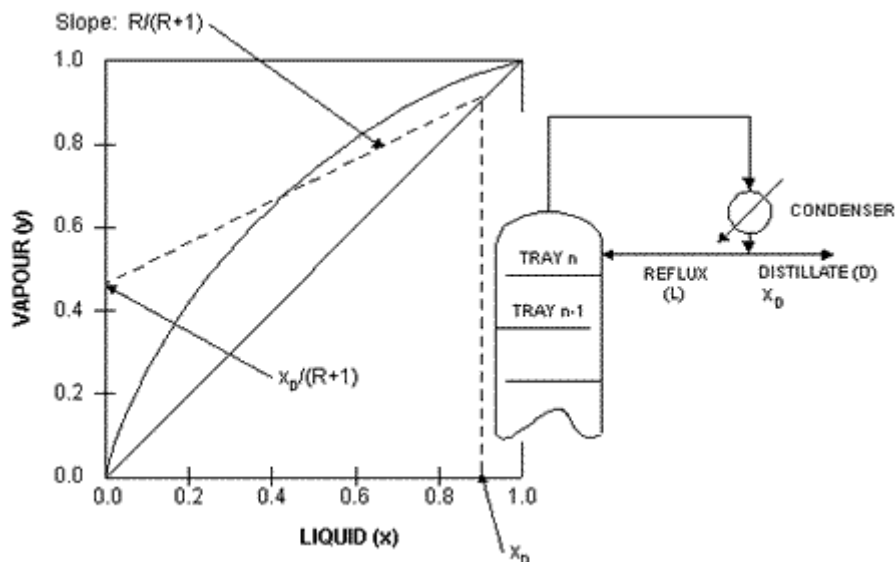
- ➔ molal heats of vaporisation of the components are roughly the same
- ➔ heat effects (heats of solution, heat losses to and from column, etc.) are negligible
- ➔ for every mole of vapour condensed, 1 mole of liquid is vaporised

The design procedure is simple. Given the VLE diagram of the binary mixture, operating lines are drawn first.

- ➔ Operating lines define the mass balance relationships between the liquid and vapour phases in the column.
- ➔ There is one operating line for the bottom (stripping) section of the column, and one for the top (rectification or enriching) section of the column.
- ➔ Use of the constant molar overflow assumption also ensures the the operating lines are straight lines.

Operating Line for the Rectification Section

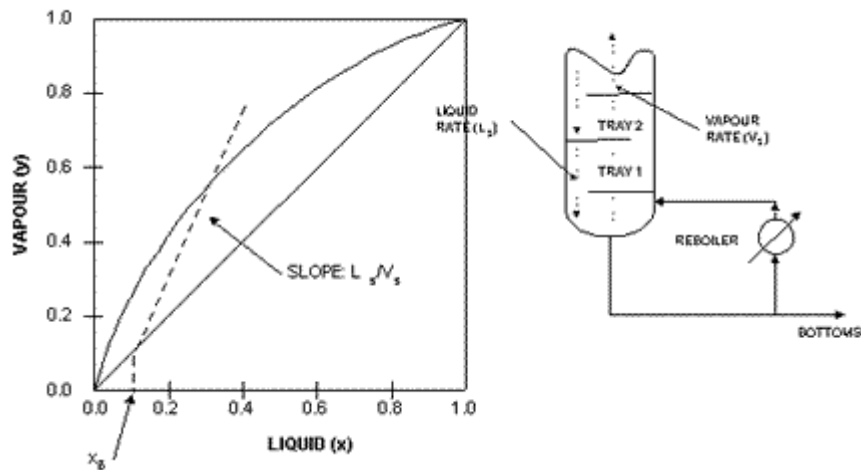
The operating line for the rectification section is constructed as follows. First the desired top product composition is located on the VLE diagram, and a vertical line produced until it intersects the diagonal line that splits the VLE plot in half. A line with slope $R/(R+1)$ is then drawn from this intersection point as shown in the diagram below.



R is the ratio of reflux flow (L) to distillate flow (D) and is called the **reflux ratio** and is a measure of how much of the material going up the top of the column is returned back to the column as reflux.

Operating Line for the Stripping Section

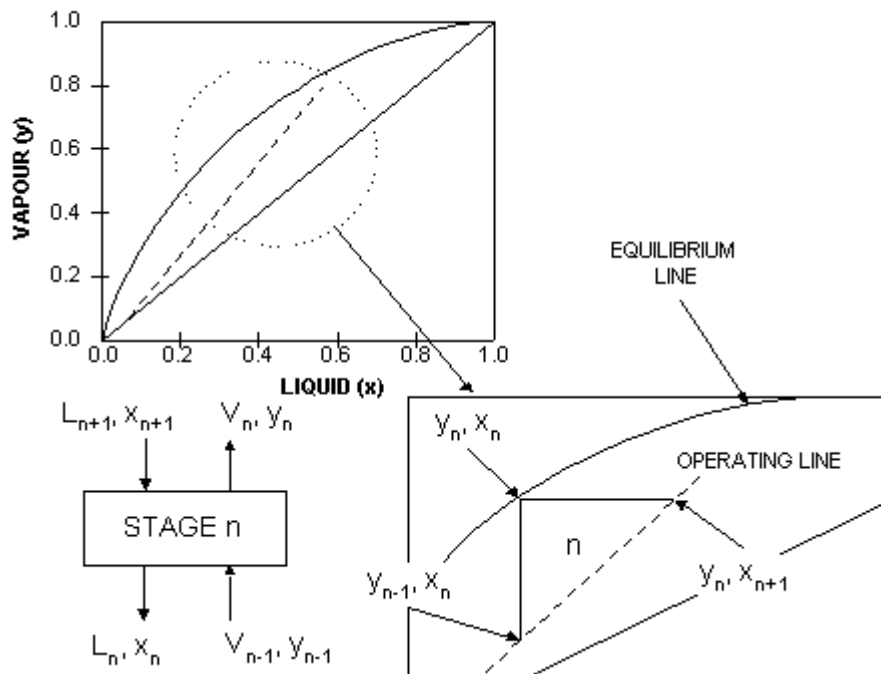
The operating line for the stripping section is constructed in a similar manner. However, the starting point is the desired bottom product composition. A vertical line is drawn from this point to the diagonal line, and a line of slope L_s/V_s is drawn as illustrated in the diagram below.



L_s is the liquid rate down the stripping section of the column, while V_s is the vapour rate up the stripping section of the column. Thus the slope of the operating line for the stripping section is a ratio between the liquid and vapour flows in that part of the column.

Equilibrium and Operating Lines

The McCabe-Thiele method assumes that the liquid on a tray and the vapour above it are in equilibrium. How this is related to the VLE plot and the operating lines is depicted graphically in the diagram on the right.

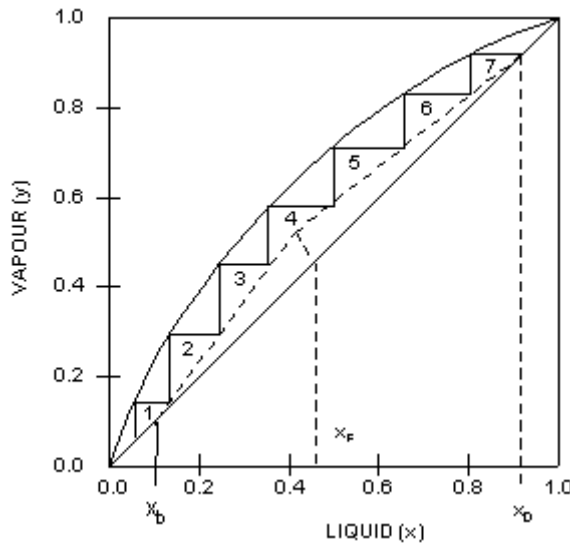


A magnified section of the operating line for the stripping section is shown in relation to the corresponding n 'th stage in the column. L 's are the liquid flows while V 's are the vapour flows. x and y denote liquid and vapour compositions and the subscripts denote the **origin** of the flows or compositions. That is ' $n-1$ ' will mean **from the stage below stage 'n'** while ' $n+1$ ' will mean **from the stage above stage 'n'**. The liquid in stage ' n ' and the vapour above it are in equilibrium,

therefore, x_n and y_n lie on the equilibrium line. Since the vapour is carried to the tray above without changing composition, this is depicted as a horizontal line on the VLE plot. Its intersection with the operating line will give the composition of the liquid on tray 'n+1' as the operating line defines the material balance on the trays. The composition of the vapour above the 'n+1' tray is obtained from the intersection of the vertical line from this point to the equilibrium line.

Number of Stages and Trays

Doing the graphical construction repeatedly will give rise to a number of 'corner' sections, and each section will be equivalent to a stage of the distillation. This is the basis of sizing distillation columns using the McCabe-Thiele graphical design methodology as shown in the following example.



Given the operating lines for both stripping and rectification sections, the graphical construction described above was applied. This particular example shows that 7 **theoretical** stages are required to achieve the desired separation. **The required number of trays (as opposed to stages) is one less than the number of stages since the graphical construction includes the contribution of the reboiler in carrying out the separation.**

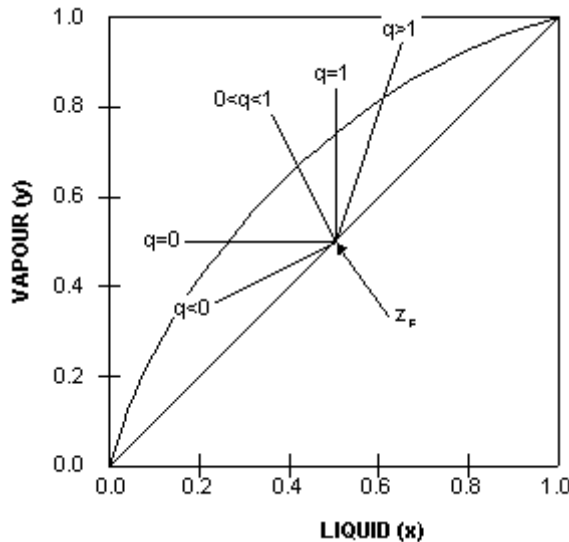
The actual number of trays required is given by the formula:

$$\text{(number of theoretical trays)} / \text{(tray efficiency)}$$

Typical values for tray efficiency ranges from 0.5 to 0.7 and depends on a number of factors, such as the [type of trays](#) being used, and internal liquid and vapour flow conditions. Sometimes, additional trays are added (up to 10%) to accommodate the possibility that the column may be under-designed.

The Feed Line (q-line)

The diagram above also shows that the binary feed should be introduced at the 4'th stage. However, if the feed composition is such that it does not coincide with the intersection of the operating lines, this means that the feed is not a saturated liquid. The condition of the feed can be deduced by the slope of the **feed line** or **q-line**. The q-line is that drawn between the intersection of the operating lines, and where the feed composition lies on the diagonal line.



Depending on the state of the feed, the feed lines will have different slopes. For example,

- $q = 0$ (saturated vapour)
- $q = 1$ (saturated liquid)
- $0 < q < 1$ (mix of liquid and vapour)
- $q > 1$ (subcooled liquid)
- $q < 0$ (superheated vapour)

The q-lines for the various feed conditions are shown in the diagram on the left.

Using Operating Lines and the Feed Line in McCabe-Thiele Design

If we have information about the condition of the feed mixture, then we can construct the q-line and use it in the McCabe-Thiele design. However, excluding the equilibrium line, only two other pairs of lines can be used in the McCabe-Thiele procedure. These are:

- feed-line and rectification section operating line
- feed-line and stripping section operating line
- stripping and rectification operating lines