

Table 6.1 Examples of industrial gas absorption processes

<i>Example</i>	<i>Industry/Purpose</i>	<i>Solvent</i>
1. Absorption of SO_3	Production of sulphuric acid and oleum	98% H_2SO_4 /Concentrated H_2SO_4
2. Absorption of H_2S	Treatment of sour natural gas	Mono- or diethanol amine
3. Drying of a gas	Air for sulphur burning	Concentrated H_2SO_4
	Drying of natural gas	Di- and tri-ethylene glycol
4. Absorption of ammonia	Ammoniation of brine in a soda ash plant	Brine
	Ammonia removal/recovery from the coke oven gas	Water or dilute H_2SO_4
5. Absorption of SO_2 from flue gases	Pollution abatement	Lime slurry
6. Absorption of NO_2	Nitric acid manufacture	Water
7. Absorption of acrylonitrile	Manufacture of acrylonitrile by ammoxidation of propylene	Water
8. Absorption of oxygen from air	Aerobic bioreactor	Fermentation broth, Waste water
9. Absorption of ethylene oxide	Ethylene oxide manufacture	Water
10. Absorption of HF	Pollution control in phosphoric acid and phosphatic fertilizer plants	Water, dilute alkali
11. Absorption of HCl gas	Hydrochloric acid production	Water
12. Absorption of CO_2	CO_2 removal from the synthesis gas in an ammonia plant	Aqueous alkanolamines, Sodium carbonate-bicarbonate buffer solution with a dissolved catalyst
13. Absorption of light hydrocarbons	Refineries	Absorption oil
14. Absorption of formaldehyde and methanol	Production of formaldehyde by air-oxidation of methanol	Water
15. Absorption of aromatics (benzene, naphthalene, etc.)	Recovery of the substances from raw coal gas	Absorption oil

recycled to the absorber. This point may be illustrated by a practical example. Absorption of carbon dioxide is an important operation in the production of synthesis gas in an ammonia plant (see Ullmann, 1985). An aqueous solution of an alkanolamine (methanolamine, ethanolamine, etc.) is commonly used. The CO₂-loaded amine solution leaving the absorption column is stripped of the dissolved CO₂ by using open steam in a 'stripping column'. The recovered CO₂ is sent to the urea plant. The regenerated or lean solvent (which may have traces of the solute, CO₂, left in it) is cooled in a heat exchanger and pumped to the amine storage tank for recycling. The arrangement is schematically shown in Figure 6.1.

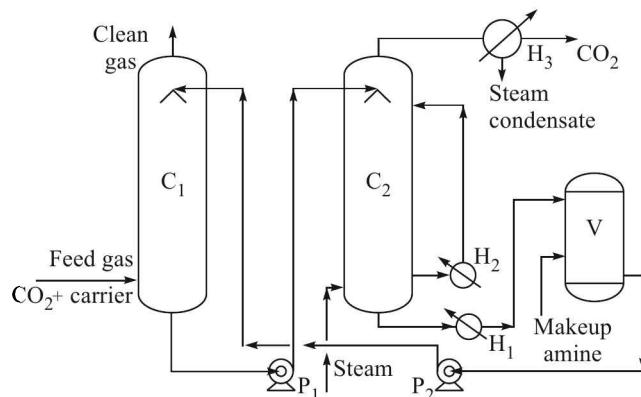


Figure 6.1 Schematic of a carbon dioxide absorption and solvent regeneration system; C₁: absorption column; C₂: regeneration column; P₁, P₂: pumps; H₁: cooler for the regenerated solvent; H₂: reboiler; H₃: condenser for steam from column C₂; V: amine storage vessel.

6.1 EQUILIBRIUM IN A GAS-LIQUID SYSTEM

A knowledge of the equilibrium distribution of a solute between two phases, i.e. the solubility of a species in a solvent, is very important in determining the driving force in any mass transfer operation. For an 'ideal' gas-liquid or vapour-liquid system, the solubility is given by Raoult's law; Henry's law is often applicable at low solute concentrations for a 'real' gas-liquid system. These laws have already been discussed in Section 4.2. The techniques of equilibrium calculation for a non-ideal solution will be discussed later.

The solubility of a gas in a liquid decreases with temperature almost as a rule. Absorption of a gas is done at a rather lower temperature so that there is a larger driving force for transfer of the solute from the gas phase to the solvent. Desorption or stripping, on the other hand, is carried out at an elevated temperature so that the direction of the driving force is reversed.

6.2 SELECTION OF SOLVENT AND STRIPPING MEDIUM

A brief but good account of the general criteria for the selection of a solvent for absorption is available in Treybal (1980). The more important criteria and considerations are described here.

(i) *Solubility:* The right solvent is one in which the species to be absorbed (i.e. the solute) is highly soluble, but it should not dissolve the carrier gas. Sometimes there may be more than one solute of similar chemical nature, but it may be desirable to remove only one of them. In such a case, a solvent which is ‘selective’ for the particular solute is preferred. For example, natural gas contains both CO_2 and H_2S as undesirable constituents. While CO_2 need not be removed when the gas is used for certain purposes, H_2S has to be. A common solvent like aqueous ethanolamine absorbs both CO_2 and H_2S . But a solution of an *hindered* amine (for example, N-methyl diethanolamine) preferentially absorbs H_2S ; only a little of CO_2 is absorbed. So it is the right solvent, although expensive, for sweetening of natural gas (Haimour et al., 1983).

As stated before, a chemical reaction between the dissolved gas and the solvent or a reagent added to it increases the rate of absorption substantially. But only if the reaction is reversible, the solute may be stripped out at a higher temperature in order to recover the solute and to regenerate the solvent. For the purpose of removal of CO_2 from the synthesis gas, aqueous caustic soda (NaOH) is *not used* as a solvent despite the high rate of absorption attainable. This is because Na_2CO_3 which is formed in solution by the reaction of CO_2 with NaOH cannot be easily decomposed to regenerate NaOH^+ . An aqueous alkanolamine is used instead since CO_2 reacts with the amine reversibly to form a carbamate. The carbamate breaks down at a higher temperature and CO_2 can be stripped out of the solution with regeneration of the solvent.



R_2NH = secondary alkanolamine (if $\text{R} = -\text{C}_2\text{H}_4\text{OH}$, it is ‘diethanolamine’); R_2NCOO^- = carbamate.

(ii) *Volatility:* A good solvent should have a low volatility or vapour pressure so that the solvent loss with the carrier gas because of vaporization remains small. A higher rate of solvent loss can cause air pollution in the nearby area, besides financial loss.

(iii) *Viscosity:* A solvent should have a low viscosity so that it flows easily. A viscous solvent used in a mechanically agitated absorber requires greater power consumption. In a packed tower, a viscous solvent causes flooding at lower liquid and gas flow rates.

(iv) *Corrosiveness:* Use of a non-corrosive or less corrosive solvent reduces the material cost of the equipment as well as its maintenance cost.

(v) *Cost:* A solvent should not be very expensive so that vaporization and other losses are not significant in terms of economy.

(vi) *Hazard and toxicity:* A solvent should not be hazardous or toxic. It should preferably be non-inflammable.

[†] On the other hand, an aqueous $\text{K}_2\text{CO}_3-\text{KHCO}_3$ buffer solution with a dissolved arsenite catalyst (Benfield process) is a common solvent for CO_2 (KHCO_3 is preferred to NaHCO_3 since the former has a higher solubility and a concentrated solution can be used for the absorption). At a lower temperature, CO_2 reacts with CO_3^{\pm} to form HCO_3^- . At an elevated temperature, HCO_3^- decomposes to release CO_2 when CO_3^{\pm} is regenerated.

The preceding criteria are general and by no means exclusive or stringent. While selecting a solvent, one has to strike a balance between the properties and efficacy of the solvent on the one hand and its cost and convenience of use on the other.

For ‘desorption’ or ‘stripping’ of a solute from the loaded solvent, a carrier gas, which is the stripping medium, has to be brought in contact with the solution. Steam is a common stripping medium. Use of steam is very convenient because the stripped solute can be easily recovered by condensing the steam leaving the stripping tower. If the solute is not soluble in water (as in the case of stripping of a hydrocarbon from an absorption oil), it can be separated from the condensate by decantation. On the other hand, if the solute is soluble in water, it can be recovered from the condensate by distillation. An inert gas like N_2 is also sometimes used as a carrier for stripping.

6.3 MINIMUM LIQUID RATE FOR ABSORPTION

Refer to Figure 4.7 in relation to gas absorption. Let G refer to the gas and L to the liquid. These symbols also represent the local gas and the liquid flow rates respectively in the column; G_s and L_s are the constant rates of flow of the two streams on *solute-free* basis (i.e. G_s and L_s are the flow rates of the ‘carrier’ phases). In gas absorption calculations, the flow rate of the feed gas G_1 , and its concentration y_1 are generally specified in a design problem. The exit gas concentration y_2 is also specified in many cases; y_2 indicates the extent of separation of the solute from the feed gas. A suitable solvent for absorption has to be selected and the equilibrium data at the chosen operating conditions (temperature and pressure) have to be obtained from the literature or from any other source.

Now the question is how to determine the required solvent flow rate for a given absorption problem. Recall the overall material balance equation (4.27) which is renumbered below.

$$G_s(Y_1 - Y_2) = L_s(X_1 - X_2) \quad (6.1)^{\dagger}$$

If the concentration of the solute in the exit gas (Y_2 , in mole ratio unit) and that in the solvent (X_2) supplied at the top of the column are known, there still remain two unknown quantities — L_s and X_1 . If one of these quantities (say L_s) is fixed, the other (X_1) can be calculated from the material balance equation given above (and vice versa). The operating line can then be drawn by locating and joining the terminal points (X_1, Y_1) and (X_2, Y_2) on the X - Y plane. The slope of the operating line is L_s/G_s . Two types of equilibrium lines and a set of operating lines of different slopes are shown in Figures 6.2(a) and 6.2(b). If L_s is reduced (i.e. a lower liquid rate is used), the slope decreases and the operating line (PQ), having one fixed terminal at the point (X_2, Y_2) , comes closer to the equilibrium line [see Figure 6.2(a)]. The driving force will decrease as a result, and a larger height of packing, or a larger number of plates (if a tray tower is used), has to be provided to achieve the desired degree of separation. But we cannot keep on reducing the liquid flow rate indefinitely. As the liquid rate is gradually reduced, the orientation of the operating

[†] The concentration of the solute in the solvent entering at the top is X_2 (or x_2 in the mole fraction unit). If the solvent is ‘fresh’, it does not contain any solute, i.e. $X_2 = 0$. However, very often the solvent is regenerated by stripping the rich solution from the absorber (see Figure 6.1) and is recycled back to the absorption tower. Traces of the solute may be left in the regenerated solvent. In such a case, $X_2 \neq 0$.

line will change and eventually a situation will arise when the operating line meets the equilibrium line at a certain point. The driving force at this point is zero. The corresponding liquid rate is the theoretical ‘minimum liquid rate’. On further reduction of the liquid rate beyond the minimum, the operating line crosses the equilibrium line which means that absorption of the gas occurs in one section of the column and desorption in the other [see the line PQ''' in Figure 6.2(b)]. This is physically impossible. So a column cannot operate, even theoretically, at a liquid rate less than the minimum as defined above.

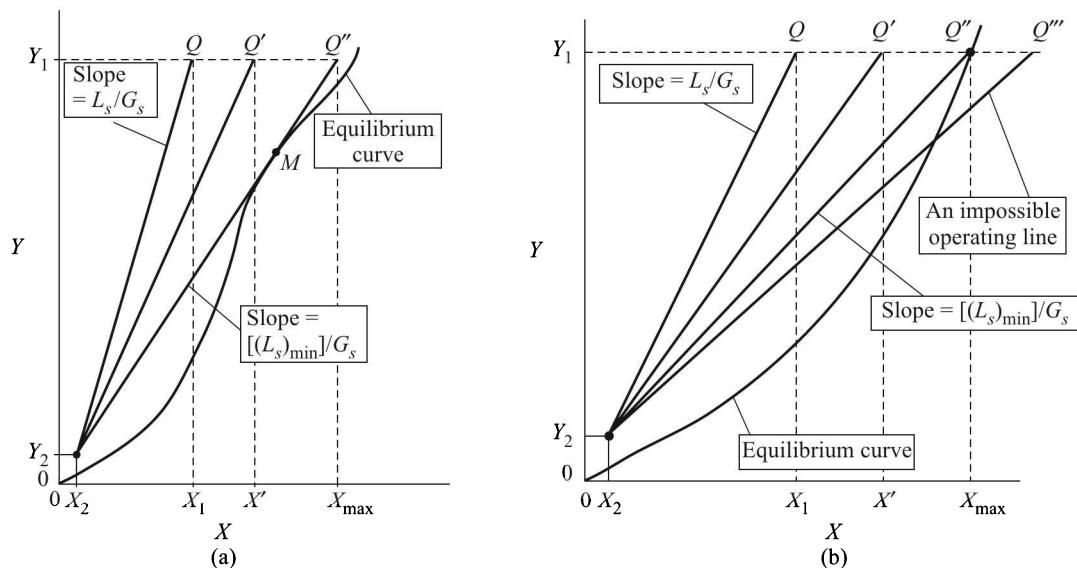


Figure 6.2(a) and (b) (a) Graphical determination of the minimum liquid rate for gas absorption (the equilibrium line is convex upwards), and (b) graphical construction for the minimum liquid rate for gas absorption (the equilibrium line is convex downwards).

The determination of the minimum liquid rate is illustrated in Figures 6.2(a) and 6.2(b) for absorption and in Figure 6.3 for stripping. In Figure 6.2(a), the equilibrium line is convex upwards around the feed gas concentration of Y_1 . The concentration pair (X , Y) at the tower bottom (for absorption) must lie on the horizontal line through Y_1 , which is the given concentration of the feed gas. As the liquid rate is decreased, the operating line changes from PQ to PQ' to PQ'' with decreasing slope. The line PQ'' is tangent to the equilibrium curve [Figure 6.2(a)] and M is the point of tangency. So

$$\text{Slope of the line } PQ'' = \frac{(L_s)_{\min}}{G_s} \quad (6.2)$$

Since G_s (the flow rate of the carrier gas) is known, $(L_s)_{\min}$ can be calculated from the slope of the line PQ'' . The corresponding concentration of the exit liquid is X_{\max} , which is the maximum possible concentration of the liquid. The driving force is zero at the point M (i.e., the point of tangency for the particular shape of the equilibrium line) which is called the *pinch point*.

The graphical construction is shown in Figure 6.2(b) for a system having an equilibrium curve which is convex downwards. In this case it is not possible to draw an operating line that will

pass through (X_2, Y_2) and touch the equilibrium line for $Y_1 > Y > Y_2$. So the minimum liquid rate is determined from the slope of the line PQ'' , Q'' being the point where the horizontal line through Y_1 meets the equilibrium curve. The driving force is zero at the point Q'' which is the *pinch point* in this case. If the liquid rate is reduced still further, we get an ‘operating line’ like PQ''' which crosses the equilibrium curve. This is physically impossible; a line like PQ''' cannot be an operating line. So PQ'' is the operating line for the theoretical minimum liquid rate.

What result do we expect if we want to find out the number of ideal plates required for the desired separation using the minimum liquid rate? In order to address this question, we redraw the equilibrium line and the operating line PQ'' for the minimum liquid in Figure 6.2(c). To find out the number of ‘ideal stages’, we may construct steps between the equilibrium line and the operating line starting from either end. Let us start at the end P . As we approach Q'' (which is the pinch point), the driving force tends to zero and it becomes impossible to reach the point Q'' by drawing a finite number of steps. Theoretically speaking, it will need an *infinite number of plates* (or an infinite height of packing, if a packed tower is used) to achieve the desired separation using the minimum liquid rate.

In the case of stripping of a solute from a solution, the feed liquid concentration X_2 is known. So Y_2 will lie on the vertical line through X_2 (Figure 6.3). The lean liquid leaves the column at

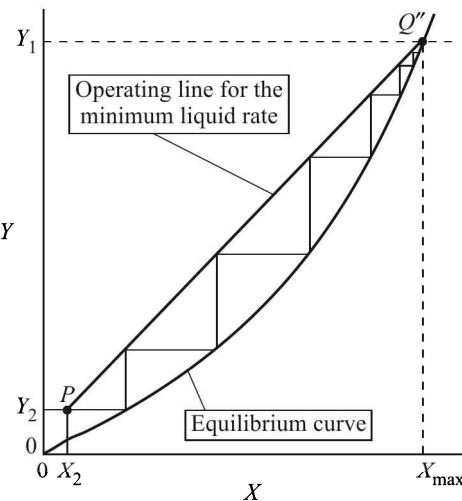


Figure 6.2(c) An infinite number of plates are required if the minimum liquid rate is used.

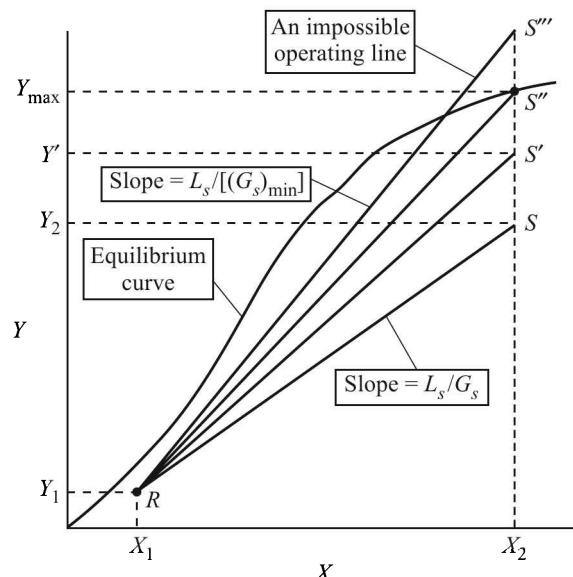


Figure 6.3 Graphical determination of the minimum gas rate for stripping of a solute.

a concentration of X_1 and the stripping gas concentration is Y_1 . The point $R(X_1, Y_1)$ represents the bottom of the tower (if the stripping gas is free from the solute, $Y_1 = 0$). Let G_s be the flow rate of the stripping gas (on solute-free basis). If we keep on reducing G_s , the orientation of the operating line changes from RS to RS' to RS'' with a gradually increasing slope. The vertical line through X_2 meets the equilibrium curve at S'' , and RS'' is the operating line for the minimum stripping gas rate, $(G_s)_{\min}$. The corresponding solute concentration in the stripping gas at the exit is the maximum, (Y_{\max}) . The driving force is zero at S'' ; it is the 'pinch point'. From the known flow rate L_s of the liquid, the minimum gas rate, $(G_s)_{\min}$, can be found out from the slope of the line RS'' . For some other shape of the equilibrium curve, it could be possible for the line RS'' to be tangent to it. In such a case, the point of tangency would be the pinch point. It all depends upon the shape of the equilibrium curve. A line like RS''' in Figure 6.3 crosses the equilibrium line and cannot be an operating line.

A few more points have to be noted in this connection. The minimum liquid rate can also be determined by a similar graphical construction on the x - y plane using the mole fraction unit of concentration. Recall that the operating line will no longer be a straight line (see Section 4.5.1)[†] in this case. The procedure described above can be followed in principle to determine the minimum liquid rate for absorption (or the minimum gas rate for stripping). But we will have to deal with two curves in order to determine the point of tangency or the pinch point. This poses a practical difficulty. So, as a matter of convenience and practice, the minimum liquid rate is determined by graphical construction on the X - Y plane as illustrated in Example 6.1.

If the equilibrium relation, expressed in terms of mole ratio unit, is linear ($Y = \alpha X$), the minimum liquid rate can be determined algebraically by calculating $X_{\max} = Y_1/\alpha$ and substituting it in the material balance Eq. (6.1). In this case, the coordinate of the pinch point is $(Y_1/\alpha, Y_1)$, and

$$[L_s]_{\min} = G_s \frac{Y_1 - Y_2}{(Y_1/\alpha) - X_2} \quad (6.3)$$

The minimum gas rate for stripping can be determined in a similar way. Also, it has to be noted that in any practical problem, the pinch point *does not* occur at the top end of the column for absorption or at the bottom end for stripping.

An absorption tower has to be operated at a liquid rate larger than the minimum. There is no general rule for determining the actual liquid rate to be used. The factors that govern the choice of the liquid rate will be discussed later. However, a liquid rate 1.2 to 2 times the minimum (Branan, 1998) is frequently used^{††}. Similar is the practice of determining the gas rate for stripping.

In this connection it will be interesting to have a closer look at the shape of an equilibrium line in connection with the determination of the minimum solvent rate for gas absorption. If Henry's law ($y = mx$) is applicable, and if $m > 1$, the equilibrium line on the X - Y plane (i.e. in the mole ratio unit) is convex downwards and the pinch point corresponding to the minimum solvent rate for absorption occurs at the bottom of the tower. On the other hand, if $m < 1$, the X - Y^* curve is convex upwards and the operating line touches the equilibrium line at some intermediate point when the minimum solvent rate is used.

[†] The equilibrium line and the operating line drawn on the x - y plane may appear to be straight lines at low concentrations of the solute.

^{††} The maximum recommended liquid rate in a packed tower varies from 40 gpm/ft² for 1-inch packing to 125 gpm/ft² for 3½-inch packing (Strigle, 1994).

EXAMPLE 6.1 (*Calculation of the minimum solvent rate*) In a petrochemical plant, a gas containing 4% *cyclo-hexane* and 96% inert has to be treated with a non-volatile absorption oil in a packed tower. It is required to remove 98% of the *cyclo-hexane* of the feed gas. The feed solvent is free from *cyclo-hexane*. If the feed gas rate is 80 kmol per hour, calculate the minimum solvent rate. The equilibrium relation is given as

$$Y = \frac{0.2X}{1 + 0.8X}$$

Solution

The following equilibrium data are calculated from the given equilibrium relation.

X	0	0.01	0.03	0.05	0.07	0.09	0.12
Y	0	0.00198	0.00586	0.0097	0.0113	0.0168	0.0219

Given: feed gas rate = 80 kmol/h; concentration of *cyclo-hexane*, $y_1 = 0.04$ (mole fraction).

Rate of input of the solute (*cyclo-hexane*) = $(80)(0.04) = 3.2$ kmol/h; carrier gas in, $G_s = 80 - 3.2 = 76.8$ kmol/h; 98% of the solute is absorbed, and 2% leaves the tower with the carrier gas.

$$Y_1 = \frac{y_1}{1 - y_1} = \frac{0.04}{1 - 0.04} = 0.0417$$

$$Y_2 = (0.02)Y_1 = 0.000834$$

Also $X_2 = 0$ (the feed solvent is solute-free). $(X_2, Y_2) \rightarrow (0, 0.000834)$

(i) Plot the equilibrium data calculated above on the X-Y plane; (ii) locate the point $Q(X_2, Y_2)$; (iii) draw the operating line through (X_2, Y_2) that touches the equilibrium line. The point of tangency P is the *pinch point*. This operating line QP corresponds to the *minimum liquid rate*; its slope is $(L_s/G_s)_{\min} = 0.19$. The graphical construction is shown in Figure 6.4.

Given: $G_s = 76.8$ kmol/h; therefore, the minimum liquid rate, $(L_s)_{\min} = (76.8)(0.19) = 14.6$ kmol/h

(The slope of QP may be obtained from the coordinates of the points P and Q as well.)

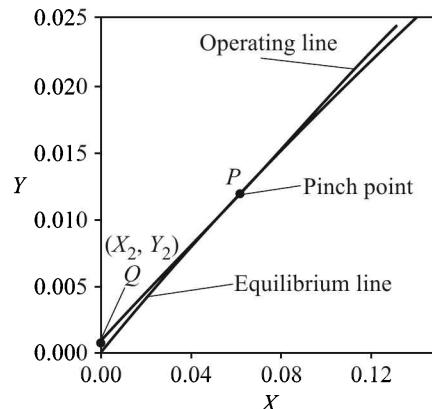


Figure 6.4 Determination of the minimum liquid rate.

6.4 DESIGN OF A PACKED TOWER

The constructional features of a packed column have already been described in the previous chapter. The packed tower and the plate tower are the two common choices for carrying out a gas absorption operation (there are other options like the venturi scrubber, the spray column, the agitated contactor, etc.). In a plate column, the gas and the liquid phases come in contact on discrete stages. So it is called ‘stage-wise contact’; and the gas and liquid concentrations undergo step changes from one plate to the next. In a packed tower, on the other hand, the liquid flowing down through the packing remains in contact with the up-flowing gas at every point of

the packed section. Also, the concentration of both phases change continuously. So, a packed column is called a ‘continuous differential contact equipment’.

Sizing of a packed column basically includes the following steps: (i) selection of the solvent; (ii) selection of packing; (iii) determination of the minimum and the actual solvent rate; (iv) determination of the column diameter; (v) determination of the packed height; and (vi) design of the liquid distributor and redistributor (if necessary), packing support and the gas distributor, design of shell, nozzles, column support, etc. (including selection of the materials to be used for the tower internals and to build the tower).

The first four steps have already been discussed. In this section, we shall concentrate upon the determination of the height of packing required for a particular separation. The discussion will be confined to the case of transfer of a single solute. Separation of multicomponent mixtures will be treated later. The following items and variables should be known or available for design purpose:

- (a) Equilibrium data
- (b) Flow rates and terminal concentrations of the gas and liquid phases
- (c) Individual or overall volumetric mass transfer coefficients, sometimes called ‘capacity coefficients’ ($k_y \bar{a}$, $k_x \bar{a}$, $K_G \bar{a}$, $K_L \bar{a}$, $K_y \bar{a}$, etc.).

6.4.1 Design Method Based on the Individual Mass Transfer Coefficients

Consider the packed tower[†] schematically shown in Figure 6.5. We use the mole fraction unit of the gas and the liquid-phase concentrations. The flow rates (G' and L') are taken on the basis of the unit cross-sectional area [i.e. mol/(time)(area)] and the specific interfacial area of contact between the gas and the liquid phases, \bar{a} , is taken on the basis of unit packed volume and has the unit of m^2/m^3 or ft^2/ft^3 . We make a steady state mass balance over a small section of the column of thickness dh .

The rate of flow of the solute (with the carrier gas) = $G'y$ mol/(time)(area).

The change in the solute flow rate over the section = $d(G'y)$; this is intrinsically negative in the case of absorption.

Let N_A be the local flux and k_y be the individual gas-phase mass transfer coefficient. Then, the packed volume in the differential section for unit cross-sectional area of the bed = $(1)(dh)$

$$\text{Interfacial area of contact in the differential section} = (\bar{a})(1)(dh)$$

$$\text{Rate of mass transfer of the solute} = (\bar{a})(dh)(N_A)$$

A mass balance over the elementary section of the bed yields

$$(\bar{a})(dh)(N_A) = -d(G'y) = -G'dy - ydG' \quad (6.4)$$

Since the carrier gas is not soluble, the change in the total gas flow rate is also equal to the rate of mass transfer of the solute, i.e.

$$-dG' = (\bar{a})(dh)(N_A) \quad (6.5)$$

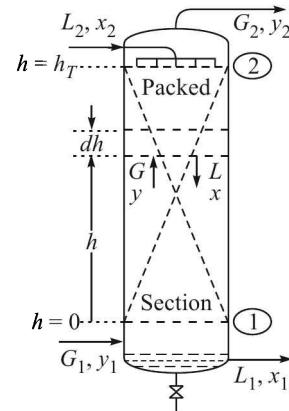


Figure 6.5 Differential section of a packed column for mass balance.

[†] As stated in Chapter 4, the bottom of the continuous contact tower is location ‘1’, and the top, location ‘2’. The flow rates and concentrations bear the corresponding subscripts.

Substituting Eq. (6.5) in Eq. (6.4), rearranging and putting $N_A = k_y(y - y_i)$,

$$(\bar{a})(dh)N_A(1 - y) = -G'dy \quad (6.6)$$

$$\Rightarrow dh = \frac{-G'dy}{k_y\bar{a}(1 - y)(y - y_i)} \quad (6.7)$$

Integrating within the appropriate limits, we get

$$h_T = \int_0^{h_T} dh = - \int_{y_1}^{y_2} \frac{G'dy}{k_y\bar{a}(1 - y)(y - y_i)} = \int_{y_2}^{y_1} \frac{G'dy}{k_y\bar{a}(1 - y)(y - y_i)} \quad (6.8)$$

Evaluation of the above integral gives the height of the packing. The integration is not straightforward, since the interfacial concentration y_i is not explicitly known as a function of the variable y . The following steps should be followed in general (McNulty, 1994):

- (a) Draw the equilibrium curve on the x - y plane for the particular gas–liquid system.
- (b) Draw the operating line from the material balance equation, see Eq. (4.26).

$$G_s \left(\frac{y}{1-y} - \frac{y_2}{1-y_2} \right) = L_s \left(\frac{x}{1-x} - \frac{x_2}{1-x_2} \right); \quad Y = \frac{y}{1+y}, \quad X = \frac{x}{1-x} \quad (6.9)$$

If the liquid mass flow rate (i.e. the rate of flow per unit cross-sectional area) is given, L_s is known. Otherwise, the minimum liquid rate on solute-free basis (L_s)_{min} is to be determined following the procedure detailed in Section 6.4. The actual liquid rate L_s is taken as a suitable multiple (commonly 1.2 to 2 times) of the minimum rate. The outlet liquid concentration x_1 is obtained from the overall material balance, Eq. (6.1).

- (c) Take any point (x, y) on the operating line (see Figure 6.6). Using the known values of k_x and k_y (or $k_x\bar{a}$ and $k_y\bar{a}$), draw a line of slope $-k_x/k_y$ from the point $S(x, y)$ to meet the equilibrium curve at $R(x_i, y_i)$ (recall Section 4.3.3). So y_i is known for the particular value of y . The line SR is called a ‘tie line’.
- (d) Repeat step (c) for a number of other points on the operating line. If k_x and k_y or their ratio are constant, a set of lines parallel to the one drawn in step (c) may be constructed. [Note that very often the mass transfer coefficients combined with the specific interfacial area (i.e. $k_x\bar{a}$ and $k_y\bar{a}$), rather than k_x and k_y , are given or known.] Now we have a set of (y, y_i) pair for $y_2 \leq y \leq y_1$.
- (e) Calculate $G = G_s(1 + y)$ at each point. Note that G_s can be calculated from the given feed gas flow rate[†].

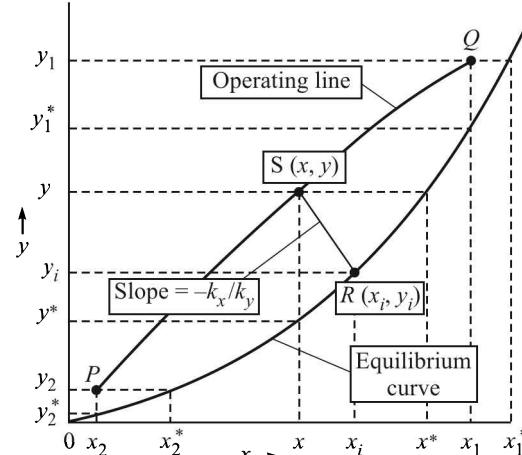


Figure 6.6 Concentration terms in the equations for packed tower design illustrated.

[†] Either the total flow rates (G , L , or G_s , L_s) or the mass flow rates (G' , L') can be used to draw the operating line. However, for the calculation of the packed height, the mass flow rates (G' , L') are used.

- (f) Calculate the value of the integrand for a set of suitably spaced values of y . Evaluate the integral in Eq. (6.8) graphically or numerically. This procedure is illustrated in Example 6.2.

The height of the packing can also be determined using other types of individual mass transfer coefficients (k_x , k_G , k_L , K_y , K_x , etc.). The design equations given below can be derived following the above procedure.

$$h_T = \int_{x_2}^{x_1} \frac{L' dx}{k_x \bar{a} (1-x)(x_i - x)} = \int_{y_2}^{y_1} \frac{G' dy}{k_G \bar{a} P (1-y)(y - y_i)} = \int_{x_2}^{x_1} \frac{L' dx}{k_L \bar{a} (C)_{av} (1-x)(x_i - x)} \quad (6.10)$$

The height of the packing for a *stripping column* can be obtained in a similar way. But here $y_2 > y_1$ and the gas-phase driving force at any point is $y_i - y$. So the design equation corresponding to Eq. (6.8) becomes

$$h_T = \int_{y_1}^{y_2} \frac{G' dy}{k_y \bar{a} (1-y)(y_i - y)} \quad (6.11)$$

6.4.2 Design Method Based on the Overall Mass Transfer Coefficient

If we express N_A in terms of the overall mass transfer coefficient [$N_A = K_y(y - y^*)$], Eq. (6.7) becomes

$$dh = \frac{-G' dy}{K_y \bar{a} (1-y)(y - y^*)} \quad (6.12)$$

Here y^* is the gas-phase concentration (in mole fraction) that is capable of remaining in equilibrium with a liquid having a bulk concentration x . The required packed height is obtained by integration of the equation between the two terminal concentrations.

$$h_T = \int_0^{h_T} dh = \int_{y_2}^{y_1} \frac{G' dy}{K_y \bar{a} (1-y)(y - y^*)} \quad (6.13)$$

Graphical or numerical integration of the right-hand side of the above equation is simpler than that of Equation (6.8). Plot the operating line, take any point (x, y) on the operating line, draw a vertical line through it and extend up to the equilibrium curve to reach the point y^* . If the values of the integrand for suitably spaced values of the variable y are calculated, the integral can be evaluated graphically or numerically.

Design equations similar to Eq. (6.10) can be obtained when the overall coefficient is given.

$$h_T = \int_{x_2}^{x_1} \frac{L' dx}{K_x \bar{a} (1-x)(x^* - x)} = \int_{y_2}^{y_1} \frac{G' dy}{K_G \bar{a} P (1-y)(y - y^*)} = \int_{x_2}^{x_1} \frac{L' dx}{K_L \bar{a} (C)_{av} (1-x)(x^* - x)} \quad (6.14)$$

The locations of x , y , y_i , y^* , x_i , x^* are schematically shown in Figure 6.6. Design equations based on the overall coefficients for a stripping operation can be easily derived from above.

6.4.3 Design Method Based on Height of a Transfer Unit

Let us recall Eq. (6.8) and write it in the following form

$$h_T = \int_{y_2}^{y_1} \frac{G' y_{iBM} dy}{k_y \bar{a} y_{iBM} (1-y)(y-y_i)} = \int_{y_2}^{y_1} \frac{G' (1-y)_{iM} dy}{k_y \bar{a} (1-y)_{iM} (1-y)(y-y_i)} \quad (6.15)$$

where y_{iBM} is the log mean value of y_B [= $(1-y)$] defined as follows:

$$y_{iBM} = (1-y)_{iM} = \frac{(1-y_i) - (1-y)}{\ln \frac{1-y_i}{1-y}} \quad (6.16)$$

[Note that we are dealing with a binary gas mixture in which B is the carrier gas (non-diffusing), and $y_B = (1-y)$; the suffix M means ‘log mean’.]

The gas-phase mass transfer coefficient often varies as $(G')^{0.8}$. Also, the ‘Colburn-Drew mass transfer coefficient’, $k'_y = k_y y_{iBM}$, remains independent of the prevailing driving force (but the coefficient k_y depends upon the concentration through y_{iBM} , see Section 3.2.2). As a result, the quantity $G'/k_y \bar{a} (1-y)_{iM}$ remains fairly constant over the packed section of the bed although the total gas mass flow rate, G' , varies. Chilton and Colburn (1935) called this quantity ‘height of a transfer unit’ based on the individual gas-phase coefficient or the ‘height of an individual gas-phase transfer unit’, denoted by H_{tG} . Taking this quantity out of the integral sign, we may rewrite Eq. (6.15) as

$$h_T = \frac{G'}{k_y \bar{a} (1-y)_{iM}} \int_{y_2}^{y_1} \frac{(1-y)_{iM} dy}{(1-y)(y-y_i)} = H_{tG} \int_{y_2}^{y_1} \frac{(1-y)_{iM} dy}{(1-y)(y-y_i)} \quad (6.17)$$

where

$$H_{tG} = \frac{G'}{k_y \bar{a} (1-y)_{iM}} = \frac{G'}{k'_y \bar{a}} \quad (6.18)$$

The integral in Eq. (6.17) is called the ‘number of individual gas-phase transfer units’, designated as N_{tG} (sometimes abbreviated NTU_G). So the packed height is

$$h_T = H_{tG} N_{tG} \quad (6.19)$$

The height of a transfer unit and the number of transfer units can be defined similarly when any other kind of mass transfer coefficient is used. For example, if the rate of mass transfer is expressed in terms of the individual liquid-phase mass transfer coefficient, the height of the packing can be written as

$$h_T = H_{tL} N_{tL} \quad (6.20)$$

where

$$H_{tL} = \frac{L'}{k_x \bar{a} x_{iBM}} = \frac{L'}{k_x \bar{a} (1-x)_{iM}}; \quad N_{tL} = \int_{x_2}^{x_1} \frac{(1-x)_{iM} dx}{(1-x)(x_i-x)} \quad (6.21)$$

and

$$(1-x)_{iM} = \frac{(1-x) - (1-x_i)}{\ln \frac{1-x}{1-x_i}} \quad (6.21)$$

Here H_{tL} is the ‘height of an individual liquid-phase transfer unit’ and N_{tL} is the ‘number of individual liquid-phase transfer units’.

If the overall gas-phase mass transfer coefficient is used to express the rate of mass transfer, the height of the packing can be obtained from the following equation:

$$h_T = \int_{y_2}^{y_1} \frac{G' y_{BM}^* dy}{K_y \bar{a} y_{BM}^* (1-y)(y-y^*)} = \frac{G'}{K_y \bar{a} y_{BM}^*} \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)} = H_{tOG} N_{tOG} \quad (6.22)$$

where

$$H_{tOG} = \text{height of an overall gas-phase transfer unit} = \frac{G'}{K_y \bar{a} y_{BM}^*}$$

$$N_{tOG} = \text{number of overall gas-phase transfer units} = \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)}$$

$$\text{and } y_{BM}^* = (1-y)_M^* = \frac{(1-y^*) - (1-y)}{\ln[(1-y^*)/(1-y)]}$$

Table 6.2 summarizes the expressions for the various forms of HTUs and NTUs.

Table 6.2 Expressions for HTUs and NTUs

Driving force	Height of a Transfer Unit (HTU)			Number of Transfer Units (NTU)	
	Symbol	DANB	ECD	Symbol	
$y - y_i$	H_{tG}	$\frac{G'}{k_y \bar{a} (1-y)_{iM}}$	$\frac{G'}{K'_y \bar{a}}$	N_{tG}	$\int_{y_2}^{y_1} \frac{(1-y)_{iM} dy}{(1-y)(y-y_i)}$
$y - y^*$	H_{tOG}	$\frac{G'}{K_y \bar{a} (1-y)_M^*}$	$\frac{G'}{K'_y \bar{a}}$	N_{tOG}	$\int_{y_2}^{y_1} \frac{(1-y)_M^* dy}{(1-y)(y-y^*)}$
$Y - Y^*$	H_{tOG}	$\frac{G'_s}{K_Y \bar{a}}$	$\frac{G'_s}{K_Y \bar{a}}$	N_{tOG}	$\int_{Y_2}^{Y_1} \frac{dY}{(Y - Y^*)}$
$x_i - x$	H_{tL}	$\frac{L'}{k_x \bar{a} (1-x)_{iM}}$	$\frac{L'}{K'_x \bar{a}}$	N_{tL}	$\int_{x_2}^{x_1} \frac{(1-x)_{iM} dx}{(1-x)(x_i - x)}$
$x^* - x$	H_{tOL}	$\frac{L'}{K_x \bar{a} (1-x)_M^*}$	$\frac{L'}{K'_x \bar{a}}$	N_{tOL}	$\int_{x_2}^{x_1} \frac{(1-x)_M^* dx}{(1-x)(x^* - x)}$
$X^* - X$	H_{tOL}	$\frac{L'_s}{K_X \bar{a}}$	$\frac{L'_s}{K_X \bar{a}}$	N_{tOL}	$\int_{X_2}^{X_1} \frac{dX}{(X^* - X)}$

$$(1-y)_{iM} = \frac{(1-y_i) - (1-y)}{\ln[(1-y_i)/(1-y)]}; \quad (1-y)_M^* = \frac{(1-y^*) - (1-y)}{\ln[(1-y^*)/(1-y)]}$$

DANB: Diffusion of A through non-diffusing B; ECD: Equimolar counterdiffusion of A and B.

Packed bed mass transfer data for gas–liquid systems are often reported in terms of the height of a transfer unit. For a particular gas–liquid system, HTU depends upon the type of packing and the gas and the liquid flow rates. The HTU data on typical systems may be obtained from the manufacturer of a particular packing.

Some qualitative physical significances can be attributed to the HTU and the NTU. The HTU indicates *inversely* the relative ease with which a given packing can accomplish separation for a particular system. For a ‘good packing’ (especially the one that provides more specific interfacial area of contact), the value of HTU is less and the packed height required for a specified degree of separation is smaller. The number of transfer units (NTU), on the other hand, indicates the difficulty of separation. The greater the extent of separation desired, the less will be the driving force available (particularly near the top of the column in case of absorption and near the bottom of the column in case of stripping), and the larger will be the NTU. A quantitative significance can be attributed to NTU in certain limiting cases. For example, in the case of absorption of a dilute gas [when $(1 - y)_M^*/(1 - y) \approx 1$], if the operating and the equilibrium lines are nearly straight and parallel, $(y - y^*)$ is approximately constant. So

$$N_{tOG} = \int_{y_2}^{y_1} \frac{(1 - y^*)_M dy}{(1 - y)(y - y^*)} \approx \int_{y_2}^{y_1} \frac{dy}{(y - y^*)} \approx \frac{y_1 - y_2}{(y - y^*)_{av}} \quad (6.23)$$

If we consider one overall gas-phase transfer unit, i.e. if we put $N_{tOG} = 1$ in the above equation, $(y_1 - y_2) \approx (y - y^*)_{av}$. Thus, a single transfer unit corresponds to the height of packing over which the change in gas concentration is approximately equal to the average driving force.

If the equilibrium relation is linear with a slope m (i.e. $y^* = mx$ or $y = mx^*$) the heights of the individual and overall transfer units are related as follows (The derivation of these equations is left as an exercise).

$$H_{tOG} = \frac{y_{iBM}}{y_{BM}^*} H_{tG} + \frac{mG'}{L'} \frac{x_{iBM}}{y_{BM}^*} H_{tL} \quad (6.24a)$$

$$\text{and} \quad H_{tOL} = \frac{L'}{mG'} \frac{y_{iBM}}{x_{BM}^*} H_{tG} + \frac{x_{iBM}}{x_{BM}^*} H_{tL} \quad (6.24b)$$

The relations may be considerably simplified if the solute concentrations are low. For example, putting

$$y_{iM} \approx y_{iM}^* \quad \text{and} \quad x_{iM} \approx x_{iM}^*,$$

$$\text{we have} \quad H_{tOG} = H_{tG} + (1/\bar{A}) H_{tL} = H_{tG} + (\bar{S}) H_{tL} \quad (6.24c)$$

where $\bar{A} = L'/mG'$ ($= L/mG$) is the absorption factor, and $\bar{S} = mG'/L'$ ($= mG/L$) is the stripping factor.

EXAMPLE 6.2 (Packed tower design using individual coefficients) A gas mixture containing 10 mol% SO₂ and 90 mol% air at 1 atm total pressure and 30°C is to be scrubbed with water to remove 97% of the SO₂ in a tower packed with 25 mm ceramic Raschig rings. The feed gas rate is 1500 kg per hour. Calculate (a) the minimum liquid rate, (b) the tower diameter if the liquid rate is 1.25 times the minimum and the tower operates at 70% of the flooding velocity, and (c) the packed height.

For the calculation of the tower diameter, the liquid may be assumed to have properties like water. Equilibrium data for SO₂-water system at 30°C may be taken from Perry's Handbook. The Colburn-Drew *volumetric mass transfer coefficients* at the given conditions are:

$$k'_x \bar{a} = 1.25 \text{ kmol}/(\text{m}^3)(\text{s})(\Delta x) \quad \text{and} \quad k'_y \bar{a} = 0.075 \text{ kmol}/(\text{m}^3)(\text{s})(\Delta y).$$

Solution

See Figure 6.5 for the notations of flow rates and concentrations.

$$\begin{aligned} \text{Average molecular weight of the feed gas (10% SO}_2, 90\% \text{ air)} &= (0.10)(64) + (0.90)(28.8) \\ &= 32.3 \end{aligned}$$

$$\text{Total feed gas rate} = 1500 \text{ kg/h} = 1500/32.3 = 46.41 \text{ kmol/h} = G_1$$

$$\text{Feed concentration, } y_1 = 0.1 \text{ (mole fraction); } Y_1 = 0.1/0.9 = 0.111 \text{ (in mole ratio unit)}$$

$$\text{Feed gas rate on solute-free basis} = G_s = G_1(1 - y_1) = (46.41)(1 - 0.10) = 41.77 \text{ kmol/h}$$

$$\text{SO}_2 \text{ entering} = G_1 y_1 = (46.41)(0.10) = 4.641 \text{ kmol/h; 97% of this SO}_2 \text{ is absorbed.}$$

$$\text{Therefore, SO}_2 \text{ absorbed} = (4.641)(0.97) = 4.502 \text{ kmol/h; SO}_2 \text{ leaving} = 0.14 \text{ kmol/h;}$$

$$\text{Concentration: } Y_2 = 0.14/G_s = 0.14/41.77 = 0.00335$$

$$y_2 = Y_2/(1 + Y_2) = 0.00335/1.00335 \approx 0.00335$$

The calculated equilibrium data in mole fraction unit (30°C, 1 atm total pressure) are given below:

$10^4 x$	0	0.562	1.403	2.8	4.22	8.42	14.03	19.65	27.9
$10^3 y$	0	0.790	2.23	6.19	10.65	25.9	47.3	68.5	104

(a) Determination of the minimum liquid rate

The equilibrium (x, y) data are converted to mole ratio units (X, Y) and plotted on the $X-Y$ plane (not shown here; this part is left as an exercise). The curve is slightly convex downwards. So the operating line corresponding to the minimum liquid rate will not touch the equilibrium line. It will rather meet the equilibrium line at the point having an ordinate $Y_1(0.111)$. This is the pinch point having the abscissa = $(X_1)_{\max} = 0.00272$. The feed water is SO₂-free. Therefore, $X_2 = 0$.

By an overall material balance,

$$\begin{aligned} G_s(Y_1 - Y_2) &= (L_s)_{\min}(X_{1,\max} - X_2) \\ \Rightarrow (41.77)(0.111 - 0.00335) &= (L_s)_{\min}(0.00272 - 0), \text{ i.e. } (L_s)_{\min} = \boxed{1655 \text{ mol/h}} \end{aligned}$$

(b) Determination of the tower diameter

The molecular weight of the solvent (water) = 18; $(L_s)_{\min} = (1655)(18) = 29,790 \text{ kg/h}$.

Take the actual liquid rate as 1.25 times the minimum. The feed liquid is *solute-free*.

The actual feed liquid rate (1.25 times the minimum),

$$L_s = (1655)(1.25) = 2069 \text{ kmol/h} (= 37,237 \text{ kg/h})$$

The liquid rate at the bottom of the tower,

$$L_1 = L_s + \text{SO}_2 \text{ absorbed} = 37,237 + (4.502)(64) = 37,525 \text{ kg/h}$$

The flooding velocity should be calculated at a point of the tower where the liquid and the gas

flow rates are maximum. In an absorption tower, it occurs at the bottom; in a stripping tower, it occurs at the top. In the given case, the flow rates at the bottom of the tower are taken.

Total pressure in the tower = 1 atm (the pressure drop across the tower is neglected); temperature = 303 K

Following the procedure of Example 5.2, the tower diameter can be calculated. At 70% of the flooding velocity, tower cross-section = 0.781 m²; diameter = 1 m; and gas mass flow rate, $G' = 1921 \text{ kg/m}^2 \cdot \text{h}$. Determination of these values is left as an exercise.

(c) Determination of the packed height

$$\text{Overall material balance, } G_s \left(\frac{y_1}{1-y_1} - \frac{y_2}{1-y_2} \right) = L_s \left(\frac{x_1}{1-x_1} - \frac{x_2}{1-x_2} \right)$$

$$G_s = 41.77 \text{ kmol/h}; L_s = 2069 \text{ kmol/h}; y_1 = 0.10; y_2 = 0.00335; x_2 = 0$$

$$\Rightarrow x_1 = 0.00217 = \text{concentration of SO}_2 \text{ in the exit solution}$$

$$X_1 = \frac{x_1}{1-x_1} = 0.002175$$

Individual gas- and liquid-phase mass transfer coefficients are given. The following equations are to be used to determine the packed height.

$$h = H_{tG}N_{tG}; \quad H_{tG} = \frac{G'}{k'_y \bar{a}} \quad \text{and} \quad N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)_{IM}}{(1-y)(y-y_i)} dy = \int_{y_2}^{y_1} f(y) dy \quad (\text{i})$$

In order to evaluate the integral (graphically or numerically), it is required to determine the interfacial concentration on the gas-side, i.e. y_i . We follow the procedure described in Section 6.4.1.

The equilibrium data are plotted on the x - y plane. The operating line as given below is also plotted on the same graph [Figure 6.7(a)].

$$G_s \left(\frac{y_1}{1-y_1} - \frac{y}{1-y} \right) = L_s \left(\frac{x_1}{1-x_1} - \frac{x}{1-x} \right)$$

$$G_s = 41.77, L_s = 2069, y_1 = 0.1, x_1 = 0.00217$$

In order to determine the interfacial gas-phase concentration y_i at a particular section of the column where the bulk concentrations are (x, y) , we proceed as follows:

(i) Locate any point $P(x, y)$ on the operating line; (ii) draw a line of slope $-k'_x \bar{a} / k'_y \bar{a}$ ($= -1.25/0.075 = -16.67$) from (x, y) to meet the equilibrium line at the point M_1 which gives the *first approximation* of the interfacial concentrations (x_i, y_i) . Because in the given system, mass transfer occurs by *diffusion of A through non-diffusing B*, the true value of (x_i, y_i) can be

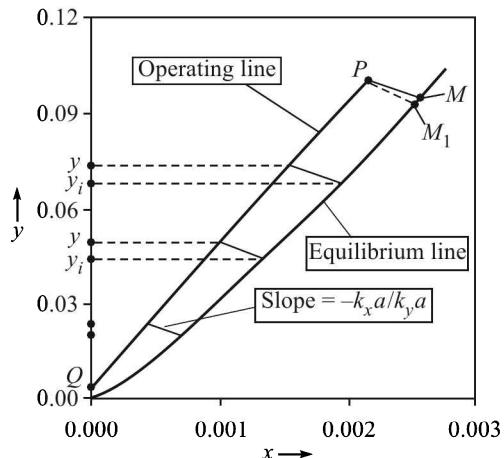


Figure 6.7(a) Determination of interfacial concentrations.

obtained by drawing a line of slope $k_x \bar{a} / k_y \bar{a}$ from the point (x, y) on the operating line to meet the equilibrium line at M_1 . Now

$$-\frac{k_x \bar{a}}{k_y \bar{a}} = -\frac{(k'_x \bar{a} / (1-x)_{iM})}{(k'_y \bar{a} / (1-y)_{iM})} = -\left(\frac{1.25}{0.075}\right)\left(\frac{(1-y)_{iM}}{(1-x)_{iM}}\right) = -16.67\left(\frac{(1-y)_{iM}}{(1-x)_{iM}}\right)$$

Using the values of interfacial concentrations at the point M_1 , the log-mean concentrations $(1-x)_{iM}$ and $(1-y)_{iM}$ can be calculated. From this we get a better estimate of $-k_x \bar{a} / k_y \bar{a}$. A line having this recalculated slope drawn from the point P meets the equilibrium line at M that gives a better estimate of (x_i, y_i) . Since we have chosen the upper terminal point $P(0.00217, 0.1)$ on the operating line and drawn a line of slope -16.67 , we reach the point $M_1(0.00255, 0.0937)$ on the equilibrium line. We have

$$(1-y)_{iM} = \frac{(1-0.0937)-(1-0.1)}{\ln \frac{1-0.0937}{1-0.1}} = 0.903$$

$$(1-x)_{iM} = \frac{(1-0.00217)-(1-0.00255)}{\ln \frac{1-0.00217}{1-0.00255}} = 0.9976$$

$$\Rightarrow -\frac{k_x \bar{a}}{k_y \bar{a}} = -(16.67)(0.903/0.9976) = -15.1$$

Draw a line of slope -15.1 through P to reach $M(0.00256, 0.0947)$ on the equilibrium line [see Figure 6.7(a)]. The point M gives a fairly accurate estimate of the interfacial concentration (x_i, y_i) . No further trial is necessary.

In a similar way, the interfacial gas-phase concentrations y_i are determined for a set of values of the bulk concentrations y . The values of the integrand, $f(y)$, are now calculated and plotted for a set of values of y .

y	0.00335	0.00887	0.0237	0.0427	0.0569	0.0734	0.1
y_i	0.00178	0.00633	0.02	0.0376	0.0513	0.068	0.047
$f(y)$	640	395	273	198	180	170	190

The area under the graph [Figure 6.7(b)] from $y_2 = 0.00335$ to $y_1 = 0.1$ is 21.5. Therefore,

$$\text{Number of gas-phase transfer units, } N_{tG} = 21.5$$

The height of a gas-phase transfer unit,

$$H_{tG} = \frac{G'}{k_y \bar{a} (1-y)_{LM}} = \frac{G'}{k'_y \bar{a}}$$

or

$$k'_y = (0.075)(3600) = 270 \text{ kmol/(h)(m}^2\text{)(}\Delta y\text{)}$$

Use the average gas flow rate to calculate H_{tG} .

Tower cross-section = 0.781 m^2

$$G'_1 = G_1 / 0.781 = 46.41 / 0.781 = 59.42 \text{ kmol/h} \cdot \text{m}^2$$

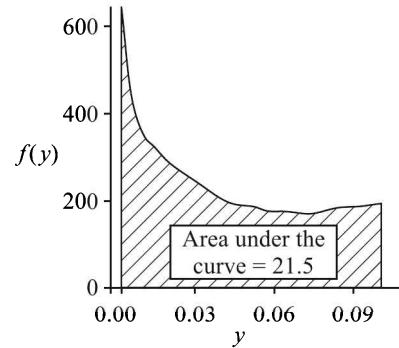


Figure 6.7(b) Graphical integration to determine N_{tG}

$$G'_2 = G_s/(1 - y_2)(0.781) = 53.66 \text{ kmol/h} \cdot \text{m}^2$$

$$G' = (G'_1 + G'_2)/2 = (59.42 + 53.66)/2 = 56.54 \text{ kmol/h} \cdot \text{m}^2$$

$$H_{tG} = 44.3/(270)(0.7823) = 0.21 \text{ m}$$

$$\text{Packed height} = H_{tG}N_{tG} = (0.21)(21.5) = \boxed{4.52 \text{ m}}$$

Note: It is easy to check that except for highly concentrated gases (> 25% of solute), it is safe

to assume that $\frac{(1-y)_{iM}}{1-y} \approx 1$, and the simplified expression for N_{tG} becomes $N_{tG} = \int_{y_2}^{y_1} \frac{dy}{y - y_i}$.

6.4.4 A Simplified Design Procedure for Dilute Gases

Making the approximation $y_{BM}^* \approx (1 - y) \approx 1$ for a dilute gas, Eq. (6.22) is reduced to

$$N_{tOG} \approx \int_{y_2}^{y_1} \frac{dy}{y - y^*} \quad (6.25)$$

If the equilibrium relation is linear, it is possible to evaluate the above integral analytically assuming that the gas and the liquid rates remain reasonably constant over the height of the packing. If L' and G' remain nearly constant (which is true for dilute gases), the equation of the operating line becomes

$$G'(y - y_2) = L'(x - x_2) \Rightarrow x = (G'/L')(y - y_2) + x_2 \quad (6.26)$$

Since $y^* = mx$ (y^* is the mole fraction of the solute in the gas phase that can remain in equilibrium with a solution having a bulk concentration x of the solute), the overall local driving force becomes

$$y - y^* = y - mx = y - [m(G'/L')(y - y_2) + mx_2] = y(1 - \bar{S}) + \bar{S}y_2 - mx_2$$

where $\bar{S} = mG'/L'$.

$$\Rightarrow N_{tOG} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} = \int_{y_2}^{y_1} \frac{dy}{y(1 - \bar{S}) + \bar{S}y_2 - mx_2} = \frac{1}{(1 - \bar{S})} \ln \frac{y_1(1 - \bar{S}) + \bar{S}y_2 - mx_2}{y_2(1 - \bar{S}) + \bar{S}y_2 - mx_2} \quad (6.27a)$$

The above equation is similar to Eq. (6.26) of McNulty (1994). It may be put in a more compact form by using the overall material balance

$$G'(y_1 - y_2) = L'(x_1 - x_2) = (L'/m)(mx_1 - mx_2) = (L'/m)(y_1^* - y_2^*)$$

$$\text{i.e. } \bar{S} = \frac{mG'}{L'} = \frac{y_1^* - y_2^*}{y_1 - y_2} \Rightarrow 1 - \bar{S} = 1 - \frac{y_1^* - y_2^*}{y_1 - y_2} = \frac{(y_1 - y_1^*) - (y_2 - y_2^*)}{y_1 - y_2}$$

$$\text{Now, } \frac{y_1(1 - \bar{S}) + \bar{S}y_2 - mx_2}{y_2(1 - \bar{S}) + \bar{S}y_2 - mx_2} = \frac{y_1 - m[(G'/L')(y_1 - y_2) + x_2]}{y_2 - mx_2} = \frac{y_1 - mx_1}{y_2 - mx_2} = \frac{y_1 - y_1^*}{y_2 - y_2^*}$$

$$\Rightarrow N_{tOG} = \frac{y_1 - y_2}{(y_1 - y_1^*) - (y_2 - y_2^*)} \ln \frac{y_1 - y_1^*}{y_2 - y_2^*} = \frac{y_1 - y_2}{(y_1 - y_1^*) - (y_2 - y_2^*)} = \frac{y_1 - y_2}{(y - y^*)_M} \quad (6.27b)$$

$$\ln \frac{y_1 - y_1^*}{y_2 - y_2^*}$$

Here $(y - y^*)_M$ is the log mean of the driving forces at the top and the bottom terminals. This is similar to 'log mean temperature difference' (LMTD) in heat transfer. (Note that $G'/L' = G/L$.)

EXAMPLE 6.3 (Tower design for scrubbing a lean gas) A stream of waste gas containing 0.6 vol% ammonia in air is to be cleaned by removing 99.5% of the ammonia before it can be discharged in the atmosphere. Because of the low concentration of the gas and the very high degree of removal desired, a dilute mineral acid is a good solvent. Dilute sulphuric acid is selected as the absorbent because of its low cost and negligible vapour pressure of the acid.

Removal of ammonia is to be done in a packed tower, 1 m² in cross-section, in which the feed gas rate is 2000 kg/h·m² and the liquid (20% H₂SO₄) is supplied at a rate of 2200 kg/h·m². The absorption is *gas-film resistance controlled* and the overall volumetric mass transfer coefficient of NH₃ on the gas-phase basis is $K_G \bar{a} = 1350 \text{ kg}/(\text{h})(\text{m}^3)(\Delta p, \text{ bar})$. Although absorption of ammonia in a dilute acid is considerably exothermic, the temperature is assumed to be constant at 28°C because of the low concentration of ammonia. The total pressure in the column is 101.3 kPa. The process starts with 1000 kg of the 20% acid solution in the storage tank for the scrubbing liquid, and the liquid is recirculated through the column continuously. The concentration of the acid in the liquid gradually decreases as a result.

- (a) How long will it take before the acid concentration falls to half of the original (in wt%)?
- (b) What total packed height is required to accomplish the separation if 0.3 m dry packing is provided above the liquid distributor in order to prevent entrainment of the liquid?
- (c) Recalculate the packed height if only 95% of the ammonia in the feed gas is removed.
- (d) What would be the value of the individual gas-phase mass transfer coefficient, k_c (kmol/(s)(m²)(ΔC)], if the effective specific interfacial area of the gas–liquid contact in the tower is estimated to be 60 m² per cubic metre of the packed volume?
- (e) Estimate the thickness of the gas-film if the diffusivity of ammonia in the carrier gas is 0.0756 m²/h.

Solution

Average molecular weight of the gas (0.6% NH₃ and 99.5% air),

$$= (0.995)(28.8) + (0.006)(17) = 28.73$$

Cross-section of the tower = 1 m²

Mass flow rate of the feed gas = 2000 kg/h·m² = 2000/28.73 = 69.61 kmol/h·m²

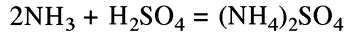
NH₃ entering with the feed = (69.61)(0.006) = 0.4177 kmol/h·m²

NH₃ absorbed = (69.61)(0.006)(0.995) = 0.4156 kmol/h·m²; NH₃ leaving = 0.00213 kmol/h·m²

Total gas leaving = 69.61 – 0.4156 = 69.19 kmol/h·m²

Mole fraction of NH₃ in the exit gas, $y_2 = 0.00213/69.19 = 0.0000308$; in the feed, $y_1 = 0.006$

(a) Amount of H_2SO_4 in the liquid at the beginning = $(1000)(0.20) = 200 \text{ kg} = 200/98 = 2.041 \text{ kmol}$



Acid consumed per hour = (moles of NH_3 absorbed)/2 = $0.4156/2 = 0.2078 \text{ kmol}$

$$\text{Time of consumption of 50% of the acid} = \frac{(2.041)(0.5)}{0.2078} = \boxed{4.91 \text{ h}}$$

(b) Assume that the operation continues till the solution remains *sufficiently acidic*. After this the solution is replaced by a fresh batch of the acid.

As the feed gas concentration is pretty low, we may take $y_{BM}^* \approx 1$; gas rate $G \approx \text{constant}$

$$H_{tOG} = \frac{G'}{K_y \bar{a} y_{BM}^*} \approx \frac{G'}{K_G \bar{a} P} = \frac{69.61 \text{ kmol/h} \cdot \text{m}^2}{\left(\frac{1350}{17}\right) \frac{\text{kmol}}{(\text{h})(\text{m}^3)(\text{bar})} (1.013)\text{bar}} = 0.865 \text{ m}$$

Again we can take $y_{BM}^*/(1 - y) \approx 1$ and $y^* = 0$ (NH_3 absorbed reacts instantaneously to form $(\text{NH}_4)_2\text{SO}_4$ in solution; the vapour pressure of NH_3 over the solution is zero). The number of transfer units is

$$N_{tOG} = \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)} = \int_{y_2}^{y_1} \frac{dy}{y} = \ln \frac{y_1}{y_2} = \ln \frac{0.006}{0.0000308} = \ln 195 = 5.27$$

$$\text{Height of packing required} = H_{tOG} N_{tOG} = (0.865)(5.27) = 4.56 \text{ m}$$

Add 0.3 m dry packing above the distributor. Total packed height = $4.56 + 0.3 = \boxed{4.86 \text{ m}}$

[If the packed height is about 20 ft or more, intermediate redistribution is necessary, and the packed height should be reasonably increased in order to accommodate the redistributor(s).]

(c) $H_{tOG} = 0.865 \text{ m}$. 95% of the NH_3 is absorbed

$$\Rightarrow N_{tOG} = \ln \frac{y_1}{y_2} = \ln \frac{0.006}{(0.006)(1-0.95)} = 3$$

Adding 0.3 m dry packing, the packed height required

$$= (H_{tOG})(N_{tOG}) + 0.3 = (0.865)(3) + 0.3 = \boxed{2.9 \text{ m}}$$

This shows that the required packed height increases by over 50% when the percentage removal of ammonia changes from 95% to 99.5%.

(d) As the absorption of NH_3 in an acid solution is gas-film diffusion controlled, the overall gas-phase mass transfer coefficient is virtually the same as the individual gas-film coefficient.

$$\Rightarrow k_G \bar{a} = K_G \bar{a} = 1350 \text{ kg/(h)(m}^3)(\Delta p, \text{ bar})$$

Specific interfacial area, $\bar{a} = 60 \text{ m}^2/\text{m}^3$; mol. wt. of $\text{NH}_3 = 17$

$$k_G = 1350/60 = 22.5 \text{ kg/(m}^2\text{)(h)(}\Delta p, \text{ bar}) = 22.5/17 = \boxed{1.324 \text{ kmol/(h)(m}^2\text{)(}\Delta p, \text{ bar})}$$

The individual gas-phase mass transfer coefficient, k_c (see Table 3.1),

$$k_c = RTk_G = (0.08317)(301)(1.324) = \boxed{33.14 \text{ kmol/(h)(m}^2\text{)(kmol/m}^3\text{)}}$$

(e) Transport of NH₃ through the gas-film occurs by ‘diffusion of A through non-diffusing B (air)’.

$$\Rightarrow k_G = \frac{D_{AB} P}{RT p_{BM} \delta}; \quad \delta = \text{film thickness}; \quad P/p_{BM} \approx 1.$$

Take $D_{AB} = 0.0756 \text{ m}^2/\text{h}$, as given.

$$\Rightarrow \delta = \frac{D_{AB}}{RT k_G} = \frac{D_{AB}}{k_c} = \frac{0.0756 \text{ m}^2/\text{h}}{33.14 \text{ (m/h)}} = 0.0023 \text{ m} = \boxed{2.3 \text{ mm}}$$

6.4.5 An Approximate Design Equation for Concentrated Gases

For a ‘concentrated gas’, a simplified form of Eq. (6.22) can be obtained by approximating the log mean quantity y_{BM}^* by the ‘arithmetic mean’.

$$\begin{aligned} y_{BM}^* &= (1 - y)_M^* \approx \frac{(1 - y^*) + (1 - y)}{2} \\ \Rightarrow N_{tOG} &= \int_{y_2}^{y_1} \frac{[(1 - y^*) + (1 - y)] dy}{2(1 - y)(y - y^*)} = \int_{y_2}^{y_1} \frac{[2(1 - y) + (y - y^*)] dy}{2(1 - y)(y - y^*)} \\ &= \int_{y_2}^{y_1} \frac{dy}{y - y^*} + \frac{1}{2} \int_{y_2}^{y_1} \frac{dy}{1 - y} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} + \frac{1}{2} \ln \frac{1 - y_2}{1 - y_1} \end{aligned} \quad (6.28)$$

The integral on the right-hand side is to be evaluated graphically or numerically. However, at the top of the column, both y and y^* are very small and the integrand will be very large. This may make the integration cumbersome. Numerical integration becomes more convenient if the integral is expressed in the form

$$\int_{y_2}^{y_1} \frac{dy}{y - y^*} = \int_{y_2}^{y_1} \frac{y - y^*}{y^*} \frac{dy}{y} = \int_{y_2}^{y_1} \frac{y}{y - y^*} d(\ln y) \quad (6.29)$$

EXAMPLE 6.4 (Packed tower design using the overall gas-phase coefficient) Ammonia is to be scrubbed from a stream of air containing 16 mole% of the solute by water at 25°C and 1 atm total pressure. The feed gas rate is 60 kmol/h·m² and the water rate is 70 kmol/h·m². It is required to remove 99% of the ammonia in the feed. If the overall gas-phase volumetric mass transfer coefficient is $K'_y \bar{a} = 100 \text{ kmol/(h)(m}^3\text{)(}\Delta y\text{)}$, determine the HTU, NTU and the height of the packing required.

$$\text{Equilibrium data: } \log \frac{P(1-x)y^*}{55.5x} = 4.699 - \frac{1922}{T}; \quad P = \text{total pressure (atm) less vapour}$$

pressure of water at the given temperature; x = mole fractions of the solute in the liquid, and y^* = equilibrium mole fraction of NH_3 in air.

Solution

The design problem involves absorption of a solute from a concentrated gas. The following calculation is based on 1 m^2 of tower cross-section and 1 hour of operation.

Molecular weight of the feed gas (16% NH_3 , 84% air) = $(0.16)(17) + (0.84)(28.8) = 26.9$

Feed gas: Feed gas rate, $G'_1 = 60 \text{ kmol/h} \cdot \text{m}^2$; mole fraction ammonia, $x_1 = 0.16$; moles ammonia input = $(60)(0.16) = 9.6 \text{ kmoles}$; moles air input = $(60)(0.84) = 50.4 \text{ kmol} (= G'_s)$; ammonia removed = 99%; moles ammonia out = $(9.6)(0.01) = 0.096 \text{ kmol}$; $G'_2 = 50.4 + 0.096 = 50.496 \text{ kmol/m}^2 \cdot \text{h}$.

$$\text{Concentration of ammonia in the outlet gas, } y_2 = \frac{0.096}{50.4 + 0.096} = 0.002$$

Equilibrium relation: The vapour pressure of water at $25^\circ\text{C} = 0.4574 \text{ psi}$; $P = 14.7 - 0.4574 = 14.24 \text{ psi} = 0.969 \text{ atm}$. Putting $P = 0.969$ and $T = 298$, the given equilibrium relation becomes

$$y^* = 1.017 \frac{x}{1-x} \quad (\text{i})$$

The liquid phase: Water rate = $70 \text{ kmol/h} (= L'_s)$; NH_3 absorbed = $(9.6)(0.99) = 9.504 \text{ kmol/h}$

$$\text{Liquid concentrations: } x_1 = \frac{9.504}{70 + 9.504} = 0.1195; \quad x_2 = 0 \text{ (since the inlet water is } \text{NH}_3\text{-free).}$$

Equation of the operating line:

$$\begin{aligned} G'_s \left(\frac{y_1}{1-y_1} - \frac{y}{1-y} \right) &= L'_s \left(\frac{x_1}{1-x_1} - \frac{x}{1-x} \right) \\ \Rightarrow (50.4) \left(\frac{0.16}{1-0.16} - \frac{y}{1-y} \right) &= (70) \left(\frac{0.1195}{1-0.1195} - \frac{x}{1-x} \right) \\ \Rightarrow \left(\frac{y}{1-y} \right) &= 1.3889 \left(\frac{x}{1-x} \right) + 0.02 \end{aligned} \quad (\text{ii})$$

We have to evaluate the integral given in Eq. (6.28) or (6.29) to determine the number of overall gas-phase transfer units. For this purpose, we have to obtain a set of values of y and y^* along the packed bed. We select a value of y ($0.002 \geq x \geq 0.16$), calculate the corresponding bulk liquid concentration x from the operating line equation (ii), and the corresponding equilibrium concentration in the gas phase using the given equation for the equilibrium relation (i) above. The following tabulated data are used for the graphical evaluation of the integral given in the alternative form, Eq. (6.29).

y	0.002	0.01578	0.02945	0.04302	0.06985	0.1093	0.1352	0.16
y^*	0.0	0.0103	0.02075	0.03145	0.0535	0.08843	0.113	0.1381

$y/(y - y^*)$	1.0	3.522	3.385	3.72	4.272	5.24	6.09	7.3
$\ln y$	-6.215	-4.15	-3.525	-3.146	-2.66	-2.214	-2.0	-1.83

The area under the plot of $y/(y - y^*)$ vs. $\ln y$ is 14.7 (see Figure 6.8).

$$\Rightarrow N_{tOG} = 14.7 + \frac{1}{2} \ln \frac{1-0.002}{1-0.16}$$

$$= 14.7 + 0.086 = \boxed{14.8}$$

(Note that the contribution of the second term is pretty small.)

Given: $K'_y \bar{a} = 100 \text{ kmol}/(\text{h})(\text{m}^2)(\Delta y)$. Average gas flow rate, $G' = (G'_1 + G'_2)/2 = 55.25 \text{ kmol}/\text{h} \cdot \text{m}^2$

$$\Rightarrow H_{tOG} = \frac{G'}{K_y \bar{a} y_{BM}^*} = \frac{G'}{K'_y \bar{a}}$$

$$= \frac{55.25}{100} = \boxed{0.552 \text{ m}}$$

Packed height = $(14.8)(0.552) = \boxed{8.2 \text{ m}}$

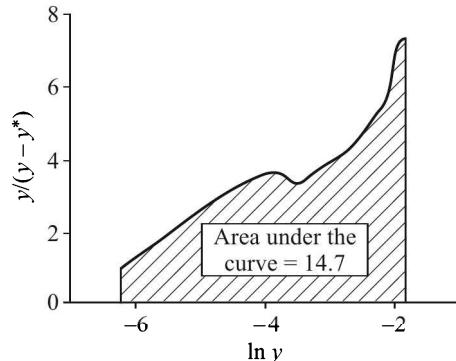


Figure 6.8 Graphical integration for the determination of N_{tOG} .

6.4.6 Analytical Determination of Number of Overall Transfer Units

If the gas and the liquid streams are ‘dilute’ and the Henry’s law ($y^* = mx$) is applicable, Eq. (6.25) can be integrated to have an algebraic equation for N_{tOG} . Substituting $x = y^*/m$ in Eq. (6.25), we get an expression for y^* in terms of y . This is substituted in Eq. (6.25) followed by integration when the following expression for N_{tOG} is obtained.

$$N_{tOG} = \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) (1 - \bar{S}) + \bar{S} \right]}{1 - \bar{S}}; \quad \bar{S} = 1/\bar{A} = mG/L \quad (6.30)$$

The above equation, suggested by Colburn in 1939, looks somewhat similar to Eq. (4.41) for calculation of the number of ideal stages. But N_{tOG} should not be confused with the number of ideal stages, although the quantities become equal if the absorption factor is unity (i.e. $\bar{A} = 1 = \bar{S}$).

6.4.7 Design Equations with Concentrations Expressed in the Mole Ratio Units

In this case the differential mass balance equation over a thin section of the bed becomes

$$G'_s dY = \bar{a} (dh) (N_A) = \bar{a} (dh) k_Y (Y - Y_i) = \bar{a} (dh) K_Y (Y - Y^*) \quad (6.31)$$

where k_Y is the individual gas-phase coefficient and K_Y is the overall gas-phase coefficient, and the concentrations are expressed in the mole ratio units. The height of the packing is given by

$$h_T = \int_{Y_2}^{Y_1} \frac{G'_s dY}{k_Y \bar{a} (Y - Y_i)} = \int_{Y_2}^{Y_1} \frac{G'_s dY}{K_Y \bar{a} (Y - Y^*)} \quad (6.32)$$

The gas flow rate on solute-free basis, G'_s , remains constant over the height of the packing. If the mass transfer coefficients also remain reasonably constant, the above equation can be written as

$$h_T = \frac{G'_s}{k_Y \bar{a}} \int_{Y_2}^{Y_1} \frac{dY}{Y - Y_i} = \frac{G'_s}{K_Y \bar{a}} \int_{Y_2}^{Y_1} \frac{dY}{Y - Y^*} \quad (6.33)$$

The design equations based on the liquid-phase mass transfer coefficients can be written following a similar procedure.

6.5 CORRELATIONS FOR MASS TRANSFER COEFFICIENTS IN PACKED TOWERS

The data on mass transfer coefficients in a packed tower are generally expressed in terms of the Sherwood number (Sh), the Colburn factor (j_D), the height of a transfer unit (H_{tG} or H_{tL}), or in the form of a direct correlation. The mass transfer coefficients (k_G , k_L , etc.) primarily depend upon (i) the type and size of the packing; (ii) the gas and the liquid flow rates; (iii) fluid properties and the transport coefficients like density, viscosity, molecular diffusivity, etc. The HTUs also depend upon these variables and parameters. The data on H_{tG} and H_{tL} for absorption of CO_2 in water (or dilute NaOH) or of NH_3 in water in a packed tower are normally available from the vendor of a packing. The former system is liquid-phase resistance controlled and is used for experimental determination of $K_L \bar{a}$ or H_{tL} ; the latter system, being gas-phase resistance controlled, is convenient for determining $K_G \bar{a}$ or H_{tG} . Typical values of the overall mass transfer coefficients in industrial columns (Strigle, 1994; Ludwig, 1997) are given in Table 6.3. Cornell et al. (1960) and Bolles and Fair (1982) reviewed and suggested correlations for the heights of gas-phase and liquid-phase transfer units for the first and second generation packings. Major correlations for HTUs and mass transfer coefficients are listed in the Perry's Handbook (7th ed., 1997, Ch. 5). Mass transfer characteristics of structured packings have been discussed by Fair and Bravo (1990), Kister (1992), Strigle (1994), Billet (1995), and others.

Table 6.3 Typical values[†] of overall mass transfer coefficients in packed towers

Packing	System	$K_G \bar{a}$ lbmol/(h)(ft ³)(atm)	System	$K_L \bar{a}$ lbmol/(h)(ft ³)(lb mol/ft ³)
#25 IMTP			$\text{CO}_2/\text{dilute NaOH}$	3.42
# 50 IMTP			$\text{CO}_2/\text{dilute NaOH}$	2.44
1-inch Pall Ring (M)			$\text{CO}_2/\text{dilute NaOH}$	3.10
#2 Superintalox (P)	$\text{NH}_3/\text{dilute acid}$	13	$\text{CO}_2/\text{dilute NaOH}$	1.5
#2 Superintalox (P)	Hydrogen chloride/Water	14	$\text{H}_2\text{S}/\text{dilute NaOH}$	4.4
1-inch Intalox saddle (C)	$\text{CO}_2/\text{dilute NaOH}$	2.82		
Intalox Structured (M)	$\text{CO}_2/\text{dilute NaOH}$	3.80		
Packing (2T)				

[†] Strigle, 1994; M: Metal; P: Plastic; C: Ceramic; 1 lbmol/(h)(ft³)(atm) = 16 kmol/(h)(m³)(atm)

A few simple correlations are cited below.

For gas-phase mass transfer in a bed with Raschig rings or Berl saddles packing, Shulman et al. (1955) suggested the following correlation.

$$j_D = 1.195 \text{Re}^{-0.36}, \quad \text{Re}_G = \frac{G' d}{(1-\varepsilon)\mu_G} \quad (6.34)$$

where G' = gas mass flow rate, μ_G = gas viscosity, ε = bed porosity, d = diameter of a sphere having the same surface area as a piece of packing.

For absorption of SO_2 in water in a packed bed with 1 inch Raschig ring packing, the following correlations for H_{tG} and H_{tL} are available.

$$H_{tG} = 1.24(G')^{0.3}/(L')^{0.25} \text{ ft} \quad (6.35)$$

$$H_{tL} = 0.37/(L')^{0.18} \text{ ft} \quad (6.36)$$

where L' , G' = liquid and gas mass flow rates, in $\text{lb}/\text{ft}^2 \cdot \text{h}$. If the values of H_{tG} and H_{tL} are available for any system (e.g. SO_2 -water system), their values for any other system can be calculated by applying the following "correction" (McNulty, 1994), provided the packing size, the packing type and the phase flow rates remain the same.

$$[H_{tG}]_1 = [H_{tG}]_2 \left[\frac{(\text{Sc})_1}{(\text{Sc})_2} \right]^{0.5}; \quad [H_{tL}]_1 = [H_{tL}]_2 \left[\frac{(\text{Sc})_1}{(\text{Sc})_2} \right]^{0.5} \quad (6.37)$$

where the suffixes 1 and 2 refer to the two systems concerned for the respective phases.

A few correlations for gas- and liquid-phase mass transfer coefficients in packed towers have been listed by Wagner et al. (1997). The correlations suggested by Onda et al. (1968) and by Billet and his group (1995) are frequently cited. Onda et al. used experimental data to develop empirical correlations for the mass transfer parameters as given below:

$$\frac{\bar{a}_{eff}}{a_p} = 1 - \exp[-1.45(\sigma_c/\sigma)^{0.75}(\text{Re}_L)^{0.1}(\text{Fr}_L)^{-0.05}(\text{We})^{0.2}] \quad (6.38a)$$

$$k_L \left(\frac{\rho_L}{\mu_L g} \right)^{1/3} = 0.0051(\text{Re}_L)^{2/3} \left(\frac{a_p}{a_{eff}} \right)^{2/3} (\text{Sc}_L)^{-1/2} (a_p d_p)^{0.4} \quad (6.38b)$$

$$k_c / (a_p D_G) = C(\text{Re}_G)^{0.7} (\text{Sc}_G)^{1/3} (a_p d_p)^{-2.0} \quad (6.38c)$$

Here $C = 2.0$ for $d_p < 15 \text{ mm}$, and $C = 5.23$ for $d_p > 15 \text{ mm}$; d_p = 'nominal packing size'; σ_c = 'critical surface tension' at the packing surface in dyne/cm (ceramics—61, polyethylene—33, glass—73, and metal—75[†]).

The correlations for k_L , k_c , the effective gas-liquid specific interfacial area of contact (\bar{a}_{eff}) and the dynamic liquid holdup (h_L) suggested by Billet and cited below are semi-empirical and are based on the 'channel model of a packed tower' (see Section 5.13) as well as on the experimental data. Billet assumed that the liquid flows as a film on the packing surface and liquid-phase mass transfer occurs according to the penetration theory over a path length l_τ (see below) with intermittent mixing of the liquid flowing down the bed.

$$\text{Liquid-phase coefficient: } k_L = (12)^{1/6} C_L \left(\frac{u_L D_L}{h_L l_\tau} \right)^{1/2} \text{ m/s} \quad (6.39a)$$

[†] The Onda et al. correlations were reassessed by Dvorak et al. (1996) for towers packed with large random packings.

Gas-phase coefficient: $k_c = C_G \left(\frac{a_p}{l_\tau} \right)^{1/2} \frac{D_G}{(\varepsilon - h_L)^{1/2}} (\text{Re}_G)^{3/4} (\text{Sc}_G)^{1/3} \text{ m/s}$ (6.39b)

Specific interfacial area: $\frac{\bar{a}_{\text{eff}}}{a_p} = 3(\varepsilon)^{0.5} (\text{Re}_L)^{-0.2} (\text{Fr}_L)^{-0.45} (\text{We}_L)^{0.75}$ (6.39c)

$$\text{Re}_G = \frac{u_G \rho_G}{a_p \mu_G}; \text{Sc}_G = \frac{\mu_G}{\rho_G D_G}; \text{Re}_L = \frac{u_L \rho_L}{a_p \mu_L}$$
 (6.40a)

$$\text{Fr}_L = \frac{u_L^2 a_p}{g}; \text{We}_L = \frac{u_L^2 \rho_L}{\sigma_L a_p}; h_L = \left(\frac{12 \text{Fr}_L}{\text{Re}_L} \right)^{1/3}$$
 (6.40b)

Here, Re = Reynolds numbers; Sc = Schmidt number; Fr = Froude number; We = Weber number; h_L = dynamic liquid holdup in the bed (volume fraction); a_p = specific surface area of the dry packing, in m^2/m^3 ; l = hydraulic diameter of the packing = $4\varepsilon/a_p$; ε = fractional bed voidage; D = diffusivity, in m^2/s ; u = superficial fluid velocity, in m/s ; subscripts L = liquid; G = gas; SI units are to be used in Eqs. 6.39(a) and 6.39(b). The constants C_L and C_G are specific to a packing. Their values for a few packings are given in Table 6.4 (for other packings, see Billet, 1995).

Table 6.4 Values* of C_L and C_G for a few random packings for use in Eqs. 6.39(a) and 6.39(b)

Packing	Material	Size	C_L	C_G
Raschig Ring	Ceramic	25 mm	1.361	0.412
Berl Saddle	Ceramic	25 mm	1.246	0.387
Pall Ring	Plastic	25 mm	0.905	0.446
Pall Ring	Metal	25 mm	1.44	0.336
Pall Ring	Metal	50 mm	1.192	0.410
NORPAC Ring	Plastic	50 mm	1.080	0.322
HiFlow Ring	Metal	50 mm	1.641	0.402

* From Billet (1995).

Recently, Piche et al. (2001) developed correlations[†] for gas- and liquid-phase mass transfer coefficients (in terms of Sherwood number) for random packings using the artificial neural network (ANN) technique^{††}. The average error in prediction is claimed to remain within 22%.

The mass transfer efficiency of a tower packing is also expressed frequently in terms of the height equivalent to a theoretical plate or stage (HETP or HETS). This is discussed in Section 6.7.

[†] The authors claim these as “confident correlations”.

^{††} Work of these authors on pressure drop in a packed bed has been referred to in Chapter 5.

EXAMPLE 6.5 It is required to design a packed tower for absorption of ethylene oxide (1.5 mol%) from a gas stream at a total pressure of 10 atm and 25°C. The feed gas rate is 15,000 kg/h and the solute concentration in the exit gas must not exceed 0.01 mol%. The solvent is ‘pure’ water. The carrier gas has properties like air and has negligible solubility in water. If 50 mm metal Pall ring is used as the tower packing, calculate the height of an overall gas-phase transfer unit using the Billet’s correlations. The solubility of ethylene oxide in water at such low concentration can be described by the Henry’s law [see Bonilla and Baron, *AIChE J.*, 1(1955) 49–54], $y = 1.07x$, at the given pressure and temperature.

The following data, information and simplifications may be used: average molecular weight of feed gas = 29.1; the operating line on the x - y plane is straight at such low concentrations; for 50 mm metal Pall ring: $\epsilon = 0.95$, $a_p = 100 \text{ m}^2/\text{m}^3$, packing factor, $F_p = 88.6 \text{ m}^{-1}$ ($= 27 \text{ ft}^2/\text{ft}^3$), $C_G = 0.41$, $C_L = 1.192$; $\mu_G = 1.8 \times 10^{-5} \text{ kg/m.s}$; $\mu_L = 0.86 \text{ cP} = 8.6 \times 10^{-4} \text{ kg/m.s}$; $\sigma_L = 72 \text{ dyne/cm}$ ($= 0.072 \text{ kg/s}^2$); $\rho_G = 11.81 \text{ kg/m}^3$ (this value is calculated by assuming the gas to be ideal); $\rho_L = 997 \text{ kg/m}^3$; diffusivities: $D_G = 1.587 \times 10^{-6} \text{ m}^2/\text{s}$, $D_L = 1.553 \times 10^{-9} \text{ m}^2/\text{s}$; the liquid rate is 1.3 times the minimum.

Solution

Only an outline of the solution is given here. The complete solution is left as an exercise.

Calculation of the water rate: Feed gas rate, $G_1 = 15,000/29.1 = 515 \text{ kmol}$; feed gas concentration, $y_1 = 0.015$; moles solute (ethylene oxide) in the feed = $(515)(0.015) = 7.8 \text{ kmol}$; moles carrier gas = $515 - 7.8 = 507.2 \text{ kmol} \approx$ moles exit gas (since the exit gas concentration is very small). Moles solute absorbed = 7.749 kmol.

Material balance to calculate the minimum liquid rate,

$$G_1 y_1 - G_2 y_2 = L_1 x_1^* - L_2 x_2$$

At the pinch point that occurs at the bottom,

$$x_1^* = y_1/1.07 = 0.015/1.07 = 0.01402; \quad y_2 \approx 0; \quad x_2 = 0.$$

$$\text{Total liquid rate at the bottom} = (515)(1.07) = 551 \text{ kmol}$$

$$\text{Min. water rate} = 551(1 - x_1^*) = 543.3 \text{ kmol}$$

$$\text{Actual water rate} = 1.3 \text{ times the minimum} = (543.3)(1.3) = 706.3 \text{ kmol}$$

$$\text{Total liquid rate at the bottom} = 706.3 + 7.749 \text{ kmol, i.e. } 13,050 \text{ kg}$$

Use this liquid rate to calculate the tower diameter.

Tower diameter: Flow parameter, $F_{lv} = (L/G)(\rho_G/\rho_L)^{0.5} = 0.0947$; taking $\Delta P/L = 0.25 \text{ inch water/ft}$ packed height, we get from Strigle’s GPDC chart (Figure 5.34),

$$\text{Ordinate} = 0.92 = C_s (F_p)^{0.5} (v_L)^{0.05}$$

Put $F_p = 27 \text{ ft}^{-1}$, $v_L = 0.86 \text{ centistoke} \Rightarrow$ capacity parameter, $C_s = (u_G)[\rho_G/(\rho_L - \rho_G)]^{0.5} = 0.178 \Rightarrow u_G = 1.626 \text{ ft/s}$. Density of the feed gas = $11.81 \text{ kg/m}^3 = 0.734 \text{ lb/ft}^3$; gas flow rate = $15,000 \text{ kg/h} = 9.186 \text{ lb/s}$

$$\Rightarrow \text{tower cross-section} = (9.186)/(0.734)(1.626) = 7.693 \text{ ft}^2$$

$$\Rightarrow \text{tower diameter} = 3.13 \text{ ft} = \boxed{0.95 \text{ m}}$$

If the pressure drop at flooding is calculated by using the Kister’s Eq. (5.23), we have $(\Delta P/L)_{\text{flood}} = 1.15 \text{ inch water/ft}$. For this pressure drop and $F_{lv} = 0.0947$, the ordinate of the GPDC chart

$= 1.5 \Rightarrow (u_G)_{\text{flood}} = 2.63 \text{ ft/s}$. This means that the column is designed at $1.626/2.63$, i.e. at 62% of the flooding velocity, which is acceptable.

Calculation of mass transfer coefficients: The relevant quantities (in SI units) are: superficial velocities, $u_G = 1.626 \text{ ft/s} = 0.496 \text{ m/s}$; $u_L = 5.087 \times 10^{-3} \text{ m/s}$ (this can be calculated from the liquid rate and the tower cross-section); $l_\tau = 4\varepsilon/a_p = (4)(0.95)/100 = 0.038$; calculate $\text{Re}_G = 3254$, $\text{Re}_L = 59$, $\text{Sc}_G = 0.9604$, $\text{Fr}_L = 2.638 \times 10^{-4}$, $\text{We}_L = 3.583 \times 10^{-3}$. Putting these values in Eqs. 6.39(a), (b) and (c)

$$h_L = 0.0377 \text{ (volume fraction)}; \quad k_L = 1.34 \times 10^{-4} \text{ m/s}$$

$$k_c = 1.485 \times 10^{-2} \text{ m/s}; \quad \bar{a}_{\text{eff}} = 77.25 \text{ m}^2/\text{m}^3$$

$$\Rightarrow k_x = (k_L \text{ m/s})(C \text{ kmol/m}^3) = (1.34 \times 10^{-4})(55.4) = 7.424 \times 10^{-3} \text{ kmol/(m}^2\text{s)(}\Delta x\text{)}$$

$$\text{and } k_y = (k_c)(P/RT) = (1.485 \times 10^{-2})[(10)/(0.0821)(298)] = 6.07 \times 10^{-3} \text{ kmol/(m}^2\text{s)(}\Delta y\text{)}$$

Heights of transfer units:

$$H_{tG} = (G')/(k'_y \bar{a}_{\text{eff}}) \approx (G')/(k_y \bar{a}_{\text{eff}}) = \boxed{0.43 \text{ m}}$$

$$H_{tL} = \boxed{0.495 \text{ m}}$$

$$\text{Stripping factor, } \bar{S} = mG/L = (1.07)(15,000/28.9)/(13,054/18) = \boxed{0.766}$$

From Eq. 6.24(c),

Height of an overall gas-phase transfer unit,

$$H_{tOG} = H_{tG} + \bar{S} H_{tL} = 0.43 + (0.766)(0.495) = \boxed{0.809 \text{ m}}$$

It is to be noted that the gas-phase and the liquid-phase resistances are comparable for this system.

6.6 DETERMINATION OF THE NUMBER OF STAGES IN A TRAY TOWER

The determination of the number of ideal stages is a major step in the design of a tray tower. In the case of absorption of a gas, the rate at which the feed enters the tower and the terminal concentrations (i.e. the concentrations of the gas at the inlet and at the exit) are specified. The minimum liquid rate is determined by following the procedure laid down in Section 6.3. The actual liquid rate L_s is usually taken as 1.2 to 2 times the minimum liquid rate. The equilibrium curve is drawn using the tabulated data or equation available and the operating line of slope L_s/G_s is drawn. The number of ideal trays is determined by drawing steps between the equilibrium curve and the operating line following the procedure given in Section 4.6.3. If the equilibrium curve is reasonably straight and has an average slope m , the number of trays can be determined algebraically using the Kremser equation, i.e. Eq. (4.41). The absorption factor, $\bar{A} = L/mG$ has been used in that equation. Similar is the procedure of graphical construction of the number of ideal stages for a stripping column. Also, the number of ideal stages for stripping can be determined algebraically from Eq. (4.44) if the equilibrium line is linear ($Y = \alpha X$)[†].

Sometimes an absorption tower is required to treat two or more feed gas streams of two different concentrations of the solute. It is *not* advisable to mix up the streams and then feed the

[†] Note that if the gas is dilute, Kremser equation [Eqs. (4.41) and (4.44)] can be used for systems having linear equilibrium relations in the form $y = mx$. The geometric value of the absorption factor should be used.

mixture to the column as this will necessitate a larger number of trays to carry out the desired separation. The feed gas streams are rather supplied to the column at different locations. For example, in gas absorption, the richest gas is fed at the bottom and the less concentrated gas stream is fed to the column at appropriate locations where the concentration of such a feed stream matches the concentration of the upcoming gas from below. Thus, if there are two feed gas streams, there will effectively be two sections of the column having two different gas flow rates, and there will be two operating lines as a result. The operating lines can be drawn using the material balance equations of the two sections separately. To determine the number of ideal trays, steps are drawn between the equilibrium curve and the two operating lines shifting from one operating line to the other as their point of intersection is crossed. This is illustrated in Example 6.7 for stripping two feed streams containing the same components, albeit at different concentrations. The procedure is also illustrated under 'Distillation' (Chapter 7).

6.7 HEIGHT EQUIVALENT TO A THEORETICAL PLATE (HETP)

The packed tower and the plate tower are the two common devices for the absorption of a solute gas from a mixture. If the former device is selected, the packed height is to be determined; in the case of a plate tower, the number of plates is to be found out for a given separation job. The ratio is the HETP [HETP = h_T/N_T ; h_T = height of packing, and N_T = number of ideal trays required to do the same separation]. Here we develop a simple expression for the HETP of a packed bed in terms of H_{tG} and H_{tL} which represent mass transfer efficiency of a packed bed.

Let us refer to Figure 6.9 in which the operating line, the equilibrium curve and the two trays (the n th and the $(n+1)$ th, represented by the two steps) are shown. Consider the $(n+1)$ th plate. The gas enters this tray at a concentration (in mole fraction) of y_{n+2} and leaves at y_{n+1} , while the liquid enters at a concentration x_n and leaves at x_{n+1} . If G' and L' are the gas and liquid flow rates per unit area at these plates (assumed to remain reasonably constant), the rate of mass transfer on the n th tray is

$$\text{Rate of mass transfer} = G'(y_{n+2} - y_{n+1}) = L'(x_{n+1} - x_n) \quad (6.41)$$

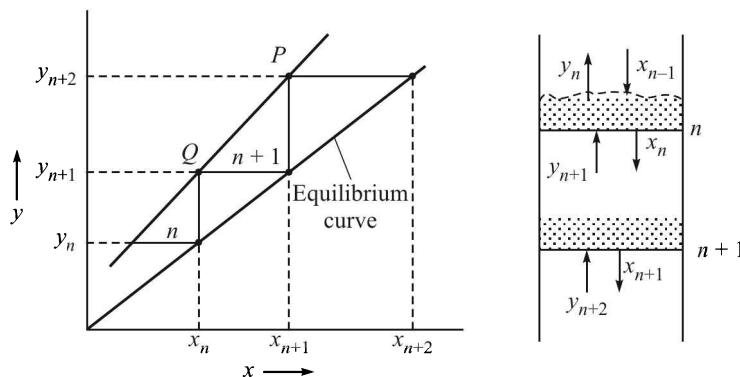


Figure 6.9 Illustration of the concept of HETP.

The rate of mass transfer over a height h of a packed section may be written as

$$\text{Rate of mass transfer} = K_G \bar{a} Ph(y - y^*)_{av} \quad (6.42)$$

where $(y - y^*)$ is the average driving force over the packed height h . If the rate of mass transfer on a tray [given in Eq. (6.41)] is the same as that achieved over a height h of a packed tower [given by Eq. (6.42)] are equal (i.e. $h = \text{HETP}$),

$$\begin{aligned} K_G \bar{a} P h (y - y^*)_{\text{av}} &= G' (y_{n+2} - y_{n+1}) \\ \text{i.e. } h &= \frac{G' (y_{n+2} - y_{n+1})}{K_G \bar{a} P (y - y^*)_{\text{av}}} \end{aligned} \quad (6.43)$$

It is reasonable to take $(y - y^*)_{\text{av}}$ as the log-mean of the overall gas-phase driving forces at the points P and Q (see Figure 6.9). Therefore,

$$(y - y^*)_{\text{av}} = \frac{(y - y^*)_P - (y - y^*)_Q}{\ln \left[\frac{(y - y^*)_P}{(y - y^*)_Q} \right]} = \frac{(y_{n+2} - y_{n+1}) - (y_{n+1} - y_n)}{\ln \left[\frac{y_{n+2} - y_{n+1}}{y_{n+1} - y_n} \right]} \quad (6.44)$$

We see from Figure 6.9,

$$\frac{y_{n+1} - y_n}{y_{n+2} - y_{n+1}} = \frac{(y_{n+1} - y_n)/(x_{n+1} - x_n)}{(y_{n+2} - y_{n+1})/(x_{n+1} - x_n)} = \frac{\text{slope of the equilibrium line}}{\text{slope of the operating line}} = \frac{m}{L'/G'} \quad (6.45)$$

$$\Rightarrow \frac{mG'}{L'} - 1 = \frac{y_{n+1} - y_n}{y_{n+2} - y_{n+1}} - 1 = \frac{(y_{n+1} - y_n) - (y_{n+2} - y_{n+1})}{y_{n+2} - y_{n+1}} \quad (6.46)$$

Comparing Eqs. (6.44), (6.45) and (6.46),

$$(y - y^*)_{\text{av}} = \frac{\left(\frac{mG'}{L'} - 1 \right) (y_{n+2} - y_{n+1})}{\ln \left(\frac{mG'}{L'} \right)} \quad (6.47)$$

Substituting for $(y - y^*)_{\text{av}}$ from Eq. (6.47) in Eq. (6.43),

$$h = \frac{G'}{K_G \bar{a} P} \frac{\ln (mG'/L')}{(mG'/L') - 1} = H_{tOG} \frac{\ln (mG'/L')}{(mG'/L') - 1} \quad (6.48)$$

But, $h =$ the packed height required to achieve the same separation as on the n th tray = HETP.

$$\Rightarrow \text{HETP} = H_{tOG} \frac{\ln (mG'/L')}{(mG'/L') - 1} = H_{tOG} \frac{\ln (\bar{S})}{\bar{S} - 1} \quad (6.49)$$

Thus, if the slope of the equilibrium line and H_{tOG} of the packing for a particular gas–liquid system are known, the HETP can be calculated. The above equation is not valid for $mG = L$.

The quantity HETP is often used to characterize the performance of a packing. A good packing has a small HETP. The concept of HETP is more widely used in packed tower distillation calculations. Its value usually ranges between 1 and 3 ft. Like HTUs, the HETP values of a packing for the separation of typical mixtures are often available from the vendor of the packing. The HETP of a packing depends upon quite a few factors (compare with HTUs) like: (i) the type and size of the packing, (ii) the gas and liquid flow rates, (iii) the physical and transport properties as well as the equilibrium relation, and (iv) the uniformity of liquid and gas distribution.

Limited data and correlations for HETP of different types of packings are available in the open literature (see, for example, Kister, 1992; Strigle, 1994; Billet, 1995; Ludwig, 1997; Perry's Handbook, 7th ed., 1997). In fact, Kister reported HETP values and charts for many of the three generations of random packings and also for structured packings. Wagner et al. (1997) reported a correlation for HETP for modern random packings. It is interesting to note that the classical data on mass transfer in a round 'wetted-wall column' reported long ago by Gilliland and Sherwood (1934) agree well with the results obtained in a bed of structured packing since mass transfer occurs in a falling film in both cases. Besides direct charts and correlations, the HETP of a packing can be calculated from the mass transfer correlations by first obtaining the HTU and then using Eq. (6.48). A few thumb rules for the estimation of HETP have been suggested by Strigle (1994).

EXAMPLE 6.6 (Tower design for scrubbing a lean gas—number of ideal stages and HETP)
Emission of toxic and hazardous volatile compounds is a challenging pollution control problem. Volatile organic compounds (VOCs) constitute the largest class of volatile emissions. An effective strategy is to construct a 'permanent total enclosure' (PTE) around the source of emission. An exhaust fan pulls out the emitted gases from the enclosure. The waste gas containing the volatiles may be cleaned by absorption using a suitable solvent or by adsorption in a solid adsorbent. The function of a PTE is to collect the emission in a rather concentrated form and to prevent its free mixing with the ambient air (mixing with air causes dilution and makes recovery almost impossible).

Carbon disulphide is a volatile hazardous substance emitted from the acid bath of a viscose plant. In a particular plant, it is planned to build a permanent total enclosure covering the acid baths. It is estimated that the exhaust gas from the PTE would leave at a rate of 12,000 Nm³ per hour containing 2% CS₂ in air. The gas has to be treated in a tower packed with 2-inch intalox saddles using an absorption oil (37 API; mol. wt. = 220; density = 840 kg/m³; viscosity = 4.7 cP at 25°C) in order to remove 98% of the CS₂. The liquid rate to be used is 1.35 times the minimum. If the overall gas-phase mass transfer coefficient is $K_G \bar{a} = 102 \text{ kmol}/(\text{h})(\text{m}^3)(\text{bar})$, calculate the diameter and height of the tower. The solution is ideal and the vapour pressure of CS₂ at 25°C is 359 mm Hg. The oil (loaded with CS₂) leaving the absorption tower is pumped to the top of a steam-stripping tower for recovery of the solute and regeneration of the solvent oil. The stripped oil having 50 ppm CS₂ is cooled and fed back to the absorption tower. The tower operates at 25°C and 1.1 bar total pressure.

Also, determine the number of ideal trays required for this separation. Note that the values of the absorption factor at the top and at the bottom of the tower differ only slightly.

Solution

The solution is ideal. Equilibrium relation,

$$y = (P^v/P)x \quad \text{where } P^v = 359 \text{ mm Hg and } P = 1.1 \text{ bar} = 827 \text{ mm Hg}$$

$$\Rightarrow y = 0.434x; \quad \text{in mole ratio unit, } Y = \frac{0.434X}{1 + 0.566X} \approx 0.434X \quad (\text{at low concentrations}) \quad (\text{i})$$

Gas flow rate and concentrations: Molecular weight of the feed (98% air, 2% CS₂)

$$= (0.98)(28.8) + (0.02)(76) = 29.7$$

Feed gas rate

$$= 12,000 \text{ m}^3/\text{h} \text{ at NTP} = 12,000/22.414 = 535.4 \text{ kmol/h} = (535.4)(29.7) = 15,901 \text{ kg/h} = G_1$$

Feed gas concentration,

$$y_1 = 0.02; \quad Y_1 = 0.02/0.98 = 0.0204$$

CS_2 entering = $(535.4)(0.02) = 10.7 \text{ kmol/h}$; air entering = $(535.4)(0.98) = 524.7 \text{ kmol/h} (= G_s)$

CS_2 absorbed = $(10.7)(0.98) = 10.5 \text{ kmol/h}$; CS_2 leaving = 0.2 kmol/h .

Total gas out = $524.7 + 0.2 = 524.9 \text{ kmol/h} = G_2$; concentration = $Y_2 = 0.2/524.7 = 0.00038$

Calculation of the minimum liquid rate: At low concentrations of the streams, the equilibrium relation in mole ratio unit can also be taken to be $Y = 0.434X$.

At the pinch point (it occurs at the bottom of the tower in this case),

$$X_1^* = Y_1/0.434 = 0.0204/0.434 = 0.047 \quad [\text{using Eq. (i)}]$$

Concentration of CS_2 in the feed solvent = 50 ppm = 50 kg CS_2 in 10^6 kg solvent

i.e. $X_2 \approx \frac{50/76}{10^6/220} = 0.000145$

Material balance: $G_s(Y_1 - Y_2) = [L_s]_{\min}(X_1^* - X_2)$

$$[L_s]_{\min} = \frac{(524.7)(0.0204 - 0.00038)}{0.047 - 0.000145} = 225.3 \text{ kmol/h}$$

Actual liquid rate (solute-free basis),

$$L_s = (1.35)[L_s]_{\min} = (1.35)(225.3) = 304.2 \text{ kmol/h}$$

Material balance: $G_s(Y_1 - Y_2) = L_s(X_1 - X_2)$

$$\Rightarrow (524.7)(0.0204 - 0.00038) = (304.2)(X_1 - 0.000145)$$

$$\Rightarrow X_1 = 0.0347 \quad \text{and} \quad x_1 = 0.0347/1.0347 = 0.0335$$

Calculation of the number of overall gas-phase transfer units:

Use the simplified equation for a dilute gas,

$$N_{tOG} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} = \frac{y_1 - y_2}{(y - y^*)_M} \quad [\text{Eq. (6.27(b))}]$$

$$y_1 = 0.02; \quad y_2 \approx 0.00038; \quad x_1 = 0.0335; \quad x_2 \approx 0.000145$$

$$y_1 - y_1^* = y_1 - mx_1 = 0.02 - (0.434)(0.0335) = 0.00546 \quad [\text{using Eq. (i)}]$$

$$y_2 - y_2^* = y_2 - mx_2 = 0.00038 - (0.434)(0.000145) = 0.000317$$

$$(y - y^*)_M = \frac{(y_1 - y_1^*) - (y_2 - y_2^*)}{\ln[(y_1 - y_1^*)/(y_2 - y_2^*)]} = \frac{0.00546 - 0.000317}{\ln(0.00546/0.000317)} = 0.001807$$

$$N_{tOG} = \frac{0.02 - 0.00038}{0.001807} = \boxed{10.8}$$

Tower diameter: Tower diameter = 3.35 m (taking the gas velocity as 65% of the flooding velocity); tower cross-section = 8.78 m^2 . (Calculations are not shown here; follow Example 6.2)

Calculation of the height of a transfer unit:

Gas flow rate at the bottom, $G_1 = 535.4 \text{ kmol/h}$; gas flow rate at the top, $G_2 = 524.9 \text{ kmol/h}$

$$\text{Average flow rate, } G_{\text{av}} = (535.4 + 524.9)/2 = 530.1 \text{ kmol/h}$$

Gas mass flow rate, $G' = (530.1 \text{ kmol/h})/(8.78 \text{ m}^2) = 60.4 \text{ kmol/h} \cdot \text{m}^2$

$$H_{tOG} = \frac{G'}{K_y \bar{a}} = \frac{G'}{K_G \bar{a} P} = \frac{60.4}{(102)(1.1)} = 0.538 \text{ m}$$

$$\text{Packed height} = (N_{tOG}) \times (H_{tOG})$$

$$= (10.8)(0.538) = \boxed{5.8 \text{ m}}$$

Number of ideal stages required: As the gas and the liquid concentrations are both low (below 5%), the equilibrium and the operating lines are practically straight. The Kremser equation can be used to determine the number of ideal stages (use the notations conforming to Figure 4.11). The average value of the absorption factor will be used.

Top of the column:

$$G_1 = 524.9 \quad \text{and} \quad L_0 = (304.2)(1 + X_0) = (304.2)(1 + 1.000145) \approx 304.2 \text{ kmol/h}$$

$$\bar{A}_1 = L_0/mG_1 = (304.2)/(0.434)(524.9) = \boxed{1.335}$$

Bottom of the column: $G_{N+1} = 535.4$; $L_N = 304.2 + 10.5 = 314.7 \text{ kmol/h}$

$$\bar{A}_N = L_N/mG_{N+1} = (314.7)/(0.434)(535.4) = \boxed{1.355}. \text{ Average value, } \bar{A} = \boxed{1.345}$$

The number of ideal stages is [use Eq. (4.41) and mole fraction unit (at low concentration)]

$$N = \frac{\log \left[\left(\frac{y_{N+1} - mx_0}{y_1 - mx_0} \right) \left(1 - \frac{1}{\bar{A}} \right) + \frac{1}{\bar{A}} \right]}{\log \bar{A}}$$

$$= \frac{\log \left[\left(\frac{0.02 - (0.434)(0.000145)}{0.00038 - (0.434)(0.000145)} \right) \left(1 - \frac{1}{1.345} \right) + \frac{1}{1.345} \right]}{\log (1.345)} = \boxed{9.5}$$

$$\text{HETP} = 5.8 \text{ m}/9.5 = \boxed{0.61 \text{ m}}$$

The absorption factor at the bottom is about 1.5% less than the value at the top. The HETP can also be calculated using Eq. (6.49) for a dilute gas–liquid system.

$$mG/L = \bar{S} = 1/1.345 = 0.7435; H_{tOG} = 0.538 \text{ m} \text{ (calculated above)}$$

$$\text{HETP} = \frac{mG}{L} \frac{\ln(mG/L)}{(mG/L) - 1} = 0.538 \frac{\ln(0.7435)}{0.7435 - 1} = \boxed{0.62 \text{ m}}$$

This is practically the same as calculated above by dividing the height of packing by the number of ideal stages.

EXAMPLE 6.7 (Design of a stripping tower with two feeds) It is required to design a stripping tower to remove *cyclo-hexane* from a non-volatile oil (molecular weight = 240). Two feed streams are available — 40 kmol/h, 10 mole% *cyclo-hexane*, and 55 kmol/h, 5.5 mole% *cyclo-hexane*. The lean oil leaving the bottom of the tower must not contain more than 0.5 mole% of the solute. The stripping column should operate at 105°C and essentially 1 atm pressure with superheated steam as the stripping medium. The steam rate is 1.2 times the minimum steam rate. Calculate the minimum steam rate and the number of ideal stages required. A solution of *cyclo-hexane* in the oil is almost ideal and Raoult's law can be used to determine the equilibrium in this system. The vapour pressure of *cyclo-hexane* can be calculated using the Antoine equation: $\ln P^v = 10.776 - (3807/T)$; P^v in bar and T in K.

Solution

The stripping column has two feeds. The more concentrated feed should enter at the top and the other feed should enter at a suitable intermediate tray of the column [Figure 6.10(a)]. The stripping steam enters the tower at the bottom. Thus the tower has two sections—the liquid rate is smaller in the upper section. But the steam rate (on solute-free basis) is constant. There will be two operating lines—the upper operating line will have a smaller slope.

Using the given equation, vapour pressure of *cyclo-hexane* at 105°C (= 378 K), $P^v = 2.023$ bar

$$\text{Equilibrium relation (Raoult's law): } Py = xP^v, P = \text{total pressure} = 1.013 \text{ bar} \Rightarrow y = 2x$$

$$\text{Equilibrium relation in the mole ratio unit: } Y = \frac{y}{1-y} = \frac{2x}{1-2x} = \frac{2(X/1+X)}{1-2(X/1+X)} = \frac{2X}{1-X}$$

The equilibrium line given by the above equation is plotted on the $X-Y$ plane, Figure 6.10(b).

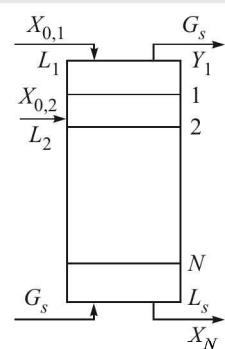


Figure 6.10(a) A stripping column with two feeds.

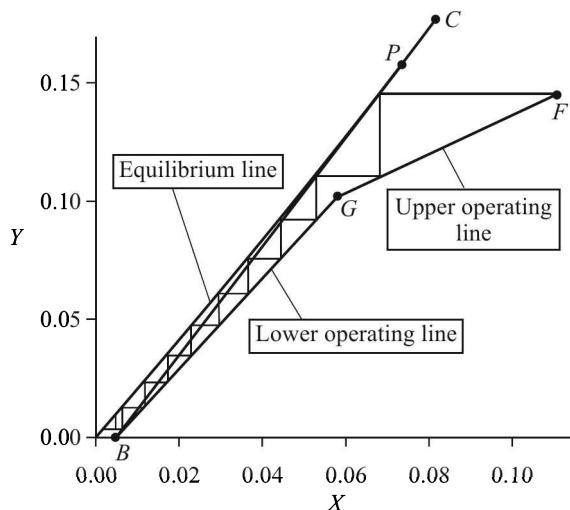


Figure 6.10(b) Graphical construction for a stripping column with two feeds.

Feed flow rates and concentrations:

Feed-1 (higher conc., to be fed at the top of the stripping tower), 40 kmol/h, 10 mole% solute.

$$L_{s1} = (40)(1 - 0.1) = 36 \text{ kmol/h; concentration (mole ratio), } X_{0,1} = 0.1/0.9 = 0.111$$

Feed-2 (lower conc., to be fed at an intermediate position in the tower), 55 kmol/h, 5.5 mole% solute

$$L_{s2} = (55)(1 - 0.055) = 52 \text{ kmol/h}$$

$$\text{concentration (mole ratio), } X_{0,2} = 0.055/(1 - 0.055) = 0.0582$$

$$\text{Total liquid in (solute-free basis), } L_s = L_{s1} + L_{s2} = 36 + 52 = 88 \text{ kmol/h}$$

$$\text{Solute concentration in the exit liquid} = 0.5 \text{ mole\%} \Rightarrow X_N = 0.005/(1 - 0.005) = 0.00502$$

No solute in the inlet stripping gas (superheated steam) $\Rightarrow Y_{N+1} = 0$.

Determination of the minimum steam rate, $G_{s,\min}$:

Locate the point $B(0.00502, 0)$ on the $X-Y$ plane [see Figure 6.10(b)].

Draw a tangent, BC , to the equilibrium line from the point B . The point of tangency, P , is the pinch point (*this has to be checked, however*).

$$\text{Slope of this tangent} = 2.316 = L_s/(G_{s,\min}) \Rightarrow G_{s,\min} = 88/2.316 = \boxed{38 \text{ kmol/h}}$$

Now we make a material balance over the column for this minimum steam rate, $G_{s,\min}$.

$$\begin{aligned} L_{s1}X_{0,1} + L_{s2}X_{0,2} + G_{s,\min}Y_{N+1} &= L_sX_N + G_{s,\min}Y_{1,\max} \\ \Rightarrow (36)(0.111) + (52)(0.0582) + (38)(0) &= (88)(0.00502) + (38)Y_{1,\max} \\ \Rightarrow Y_{1,\max} &= 0.1732 \end{aligned}$$

Slope of the operating line of the upper section for this steam rate $= L_s/G_{s,\min} = 36/38 = 0.9474$. In order to check if the point P is the *true* pinch point, the following steps are to be followed.

- (i) Locate the point $D(X_{0,1}, Y_{1,\max})$, i.e. $(0.111, 0.1732)$. Draw a line of slope 0.9474 through D . This line meets BC at E [not shown in Figure 6.10(b)].
- (ii) If the point E lies above P , then P is the true pinch point. If not, a smaller gas rate is selected and the two operating lines are drawn. The gas rate for which the two operating lines meet on the equilibrium line, is the true minimum stripping gas rate. This gas rate is to be found by trial-and-error. However, in the given problem the equilibrium line is only slightly curved and the above exercise does not yield an appreciably different gas rate than that calculated from the slope of the line BC . So we take $G_{s,\min} = \boxed{38 \text{ kmol/h}}$.

The operating lines: Actual steam rate (1.2 times the minimum), $G_s = (1.2)(38) = 45.6 \text{ kmol/h}$. The overall material balance is made again to calculate the actual solute concentration Y_1 in the steam leaving at the top.

$$\begin{aligned} L_{s1}X_{0,1} + L_{s2}X_{0,2} + G_sY_{N+1} &= L_sX_N + G_sY_1 \\ \Rightarrow (36)(0.111) + (52)(0.0582) + (45.6)(0) &= (88)(0.00502) + (45.6)Y_1 \\ \Rightarrow Y_1 &= 0.1443 \end{aligned}$$

$$\text{Slope of the upper operating line} = L_s/G_s = 36/45.6 = 0.7895$$

$$\text{Slope of the lower operating line} = L_s/G_s = 88/45.6 = 1.93$$

Draw the lower operating line, BG , through the point B having the slope 1.93. Locate the point F (0.111, 0.1443) corresponding to the top of the column, draw the upper operating line, FG , through F having the slope 0.7895. The two operating lines meet at the point G [Figure 6.10(b)].

Number of ideal stages: The number of ideal stages can be determined by drawing steps between the operating lines and the equilibrium line. We start from the point F , draw steps changing from the upper to the lower operating line as the common point G is crossed. A total of 9.4 ideal plates are required. The second feed enters the tower at the second plate from the top where we move from the upper to the lower operating line during construction of steps.

(Note: Construction of steps may start from the lower point B as well.)

6.8 TRAY EFFICIENCY

In Chapter 5, we introduced the concept of an ideal stage. An *ideal stage* provides quite an efficient contact between the phases so that they attain equilibrium. Or, in other words, the phases leaving an ideal stage are at equilibrium irrespective of the inlet concentrations. However, the performance of a *real stage* will expectedly be different from that of an ideal stage. The *tray efficiency* is an indicator of how closely the performance of a real tray approaches that of an ideal tray.

6.8.1 Point Efficiency

The contact between the phases on a tray is not uniformly good at all locations, i.e. the tray efficiency is likely to vary from one location to another on a tray. However, if we assume that (i) the liquid is ‘vertically well-mixed’ (i.e. the liquid concentration is uniform in the vertical direction at any point on the tray and (ii) the gas-phase is in plug flow (i.e. the gas concentration changes along the depth), we can define a ‘point efficiency’ based on the gas-phase concentration. In the following analysis, we consider gas–liquid contact on a tray, but the results are applicable to other types of systems as well. The point efficiency is defined as

$$E_{OG} = \frac{y_{n,\text{local}} - y_{n+1,\text{local}}}{y_{n,\text{local}}^* - y_{n+1,\text{local}}} \quad (6.50)$$

The notations used in Eq. (6.50) are illustrated in Figure 6.11. The local concentration (i.e. the concentration at any point or location on the tray) of the gas leaving the $(n+1)$ th tray is $y_{n+1,\text{local}}$. This is also the concentration of the gas entering the n th tray. Similarly, $y_{n,\text{local}}$ is the local concentration of the gas leaving the n th tray. The local concentration of the liquid is $x_{n,\text{local}}$ (note that this concentration is the same at all depths at a given location on the tray since the liquid is assumed to be ‘vertically well-mixed’), and the concentration of the solute in the gas phase that can remain in equilibrium with a liquid of this concentration is $y_{n,\text{local}}^*$. So E_{OG} is the ratio of actual enrichment of the gas (as it bubbles through the liquid on the tray) to the maximum enrichment that could have been achieved if the exiting gas had reached equilibrium with the local liquid.

Tray efficiency depends upon the rate of interphase mass transfer in the dispersion on a tray. If the rate of mass transfer is very high, the efficiency will be close to 100%. However, at a high

rate of mass transfer, the phases will approach equilibrium and the mass transfer driving force will be rather small. Therefore, a large mass transfer coefficient and a large interfacial area are required in order to achieve a high tray efficiency. We shall show below how the point efficiency is related to these quantities.

Let us refer to Figure 6.11(b) in which a small area dS of the tray and the gas–liquid dispersion over it are shown. Let G' be the molar gas flow rate (on unit area basis) which is assumed to remain constant. We write the steady-state mass balance over a small thickness dz of the dispersion.

$$\text{The rate of mass transfer to the gas phase} = G'dS dy \quad (6.51)$$

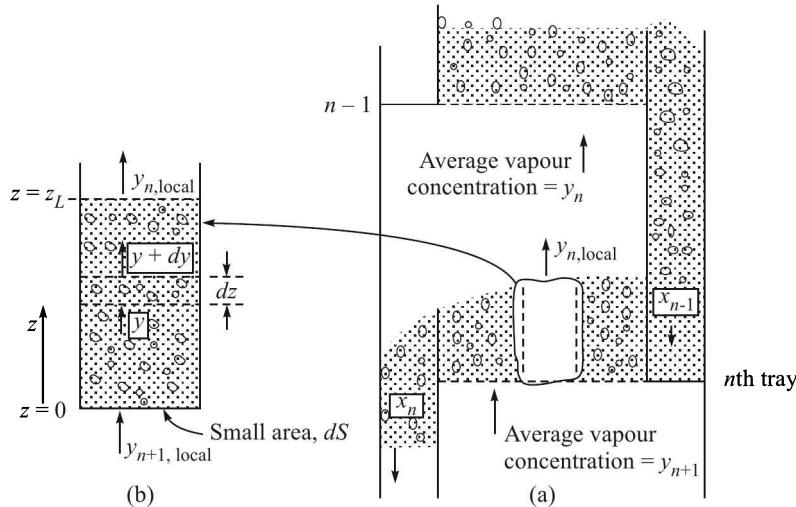


Figure 6.11 (a) Point efficiency illustrated; (b) local differential mass balance over a small element of the gas–liquid dispersion.

The rate of mass transfer (it is assumed that mass transfer occurs from the liquid to the gas phase)

$$= K_y(\bar{a} dS dz)(y_{n,\text{local}}^* - y) \quad (6.52)$$

where

K_y = overall gas-phase mass transfer coefficient

\bar{a} = specific interfacial area (say square metre of gas–liquid contact area per cubic metre volume of the dispersion)

$dS.dz$ = small volume of the dispersion

$\bar{a} dS dz$ = gas–liquid contact area in the small volume of the dispersion

$(y_{n,\text{local}}^* - y)$ = local mass transfer driving force at a height z from the tray floor.

Equating the right-hand sides of Eqs. (6.51) and (6.52),

$$G'dS dy = K_y(\bar{a} dS dz)(y_{n,\text{local}}^* - y)$$

Integrating over the height z_L of the dispersion [see Figure 6.11(b)],

$$\int_{y_{n+1,\text{local}}}^{y_{n,\text{local}}} \frac{dy}{(y_{n,\text{local}}^* - y)} = \int_0^{z_L} K_y \bar{a} \frac{dz}{G'} \quad (6.53)$$

Since the local liquid is vertically well-mixed, x_{local} is uniform over the depth z_L of the dispersion. So the corresponding equilibrium gas-phase concentration, $y_{n,\text{local}}^*$ is also constant. So we have from Eq. (6.53),

$$-\ln \frac{y_{n,\text{local}}^* - y_{n,\text{local}}}{y_{n,\text{local}}^* - y_{n+1,\text{local}}} = \frac{K_y \bar{a} z_L}{G'} = N_{tOG} \quad (6.54)$$

$$\Rightarrow -\ln \left[1 - \frac{y_{n,\text{local}} - y_{n+1,\text{local}}}{y_{n,\text{local}}^* - y_{n+1,\text{local}}} \right] = -\ln(1 - E_{OG}) = N_{tOG}$$

$$\Rightarrow E_{OG} = 1 - e^{-N_{tOG}} \quad (6.55)$$

Here N_{tOG} is the number of overall gas-phase transfer units. A similar expression based on liquid-phase mass transfer can be easily derived. If the number of transfer units is high (which is possible if $K_y \bar{a}$ is high), then E_{OG} approaches unity, i.e. the tray approaches ideal performance.

EXAMPLE 6.8 (Point efficiency on a tray) For a particular tray in a gas absorption column, the following data are available: superficial gas flow rate (taken on the basis of ‘active tray area’), $G' = 198 \text{ kmol/m}^2\cdot\text{h}$; overall gas-phase volumetric mass transfer coefficient, $K_c \bar{a} = 2.1 \text{ kmol}/(\text{m}^3)(\text{s})(\text{kmol}/\text{m}^3)$; depth of the froth on the tray, $z_L = 0.25 \text{ m}$; total pressure, $P = 1.2 \text{ bar}$; temperature, $T = 35^\circ\text{C}$. Calculate the point efficiency.

Solution

Calculate the overall mass transfer coefficient, $K_y \bar{a}$. Given, $T = 308 \text{ K}$.

$$K_y \bar{a} = K_c \bar{a} (P/RT) = (2.1)[(1.2)/(0.08317)(308)] = 0.0984 \text{ kmol}/(\text{s})(\text{m}^3)(\Delta y)$$

From Eq. (6.54),

$$N_{tOG} = \frac{K_y \bar{a} z_L}{G'} = \frac{(0.0984)(0.25)}{(198/3600)} = 0.447$$

Point efficiency [Eq. (6.55)],

$$E_{OG} = 1 - e^{-N_{tOG}} = 1 - e^{-0.447} = 1 - 0.64 = \boxed{0.36}$$

6.8.2 Murphree Efficiency

It is quite reasonable to assume that the gases leaving the dispersion at different locations of a tray get mixed up before entering the upper tray. The concentration of the liquid on a tray changes as it flows over the tray. Let x_n be the concentration of the liquid leaving the n th tray and y_n^* be the corresponding equilibrium concentration of the gas phase. On the basis of these concentrations, Murphree (1925) defined a tray efficiency as given below.

$$E_{MG} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} \quad (6.56)$$

Here y_n and y_{n+1} are the average concentrations of gases leaving the n th and the $(n+1)$ th tray respectively [see Figure 6.11(a)].

The Murphree tray efficiency can be related to the ‘point efficiency’ depending upon the extent of mixing of the liquid as it flows across a tray as well as of the vapour between two consecutive trays. The following limiting cases may be identified.

(a) If the liquid phase is assumed to be ‘well-mixed’ over a tray, the liquid composition remains uniform at x_n . Then

$$E_{MG} = E_{OG} \quad (6.57)$$

The above assumption is good for a tower of small diameter. In such a tower, the liquid moves through a rather small distance on a tray to reach the downcomer. The gas bubbling through the liquid creates enough turbulence to ensure almost uniform concentration of the liquid over the tray. However, in a tower of larger diameter, the liquid has to travel through a longer distance to reach the downcomer and its concentration varies substantially from one end of the tray to the other (i.e. along its flow path). The extent of ‘backmixing’ on a tray is generally taken into account by introducing an ‘axial dispersion coefficient’ (see the Appendix).

(b) The other limiting case arises if the liquid is in ‘plug flow’ and no axial mixing at all occurs in the liquid. In such a case, the following relation between the Murphree efficiency and the ‘point efficiency’ can be developed (this is called ‘Lewis case 1’).

$$E_{MG} = \frac{L'}{mG'} \left[\exp\left(\frac{mG'E_{OG}}{L'}\right) - 1 \right] \quad (6.58)$$

Here L' = liquid mass flow rate in mole/(area)(time) and m = Henry’s Law constant [$y_n^* = mx_n$].

EXAMPLE 6.9 (Murphree tray efficiency) The following data were collected by analyzing the gas and liquid samples to and from a particular tray (call it the n th tray) in an experiment on plate efficiency in an absorption column.

$$G_s = 90 \text{ kmol/h}; \quad L_s = 100 \text{ kmol/h}; \quad y_{n+1} = 0.12; \quad x_n = 0.078; \quad x_{n-1} = 0.06$$

The equilibrium relation is linear in the form, $y = 1.01x$. If the liquid on the tray is ‘well mixed’, calculate the Murphree tray efficiency.

Solution

First we calculate y_n from the material balance equation.

$$\begin{aligned} X_n &= x_n/(1 - x_n) = 0.078/0.922 = 0.0846. \text{ Similarly, } X_{n-1} = 0.06383; Y_{n+1} = 0.1364. \\ G_s(Y_{n+1} - Y_n) &= L_s(X_n - X_{n-1}) \Rightarrow (90)(0.1364 - Y_n) = (100)(0.0846 - 0.06383) \\ \Rightarrow Y_n &= 0.1133 \\ y_n &= Y_n/(1 + Y_n) = 0.1133/1.1133 = 0.1018 \\ y_n^* &= mx_n = (1.01)(0.078) = 0.0788 \\ \text{Murphree efficiency, } E_{MG} &= \frac{y_{n+1} - y_n}{y_{n+1} - y_n^*} = \frac{0.12 - 0.1018}{0.12 - 0.0788} = 0.442, \text{ i.e. } 44.2\% \end{aligned}$$

EXAMPLE 6.10 (*Murphree efficiency when the liquid is in plug flow*). If, in Example 6.8, the liquid is in ‘plug flow’ and its flow rate L' is $240 \text{ kmol/m}^2 \cdot \text{h}$, the equilibrium relation is $y = 1.01x$, calculate the Murphree efficiency.

Solution

The absorption factor,

$$\bar{A} = L'/mG' = (240)/(1.01)(198) = 1.2$$

Using Eq. (6.58),

$$\text{Murphree efficiency, } E_{MG} = \bar{A} [\exp(E_{OG}/\bar{A}) - 1] = 1.2[\exp(0.36/1.2) - 1] = 0.42, \text{ i.e. } \boxed{42\%}$$

6.8.3 Effect of Liquid Entrainment on Tray Efficiency

Entrainment of liquid droplets in the up-flowing gas considerably affects the efficiency. If E is the fractional entrainment (moles of liquid entrained per mole of liquid entering a tray), the Murphree efficiency, corrected for the entrainment, is given by

$$E_{MGE} = \frac{E_{MG}}{1 + E_{MG}[E/(1 - E)]} \quad (6.59)$$

6.8.4 Overall Tray Efficiency

An *overall tray efficiency* (or overall column efficiency in a section of the column) is sometimes used to determine the number of real trays. It is defined as

$$E_O = \frac{\text{number of ideal trays}}{\text{number of real trays}} \quad (6.60)$$

The overall tray efficiency E_O is sometimes obtained from plant data. If (i) the gas and liquid flow rates (G', L') and the equilibrium parameter m remain constant over a section of the column, and (ii) the Murphree efficiency remains the same for all the trays, the following relation between the overall column efficiency and the Murphree efficiency can be derived.

$$E_O = \frac{\ln \left[1 + E_{MG} \left(\frac{mG'}{L'} - 1 \right) \right]}{\ln \left(\frac{mG'}{L'} \right)} \quad (6.61)$$

If $\frac{mG'}{L'} \sim 1$, E_O and E_{MG} become nearly equal.

EXAMPLE 6.11 (*Determination of the number of real trays if the Murphree efficiency is known*) A gas absorption column is to receive 130 kmol/h of feed gas containing 12 mole% of a solute. It is required to remove 93% of the solute using 150 kmol/h of a solvent. The feed solvent has 0.4 mol% of residual solute in it. The Murphree tray efficiency is known to be 0.45. Determine the number of ideal trays as well as the number of real trays required for the separation. The equilibrium data for the system are given as:

x	0.0133	0.0333	0.0493	0.064	0.0747	0.0933	0.1053
y	0.01	0.0266	0.0433	0.06	0.0733	0.1	0.12

Solution*Flow rates:*

$$G_s = (130)(1 - 0.12) = 114.4; \quad L_s = (150)(1 - 0.004) = 149.4 \text{ kmol/h}$$

Concentrations: feed liquid: $x_0 = 0.004, X_0 = 0.004/0.996 = 0.00402$; feed gas: $y_{N+1} = 0.12, Y_{N+1} = 0.12/0.88 = 0.1364$; exit gas (93% of the solute is removed): $Y_1 = (0.1364)(1 - 0.93) = 0.00945$.

Material balance: $G_s(Y_{N+1} - Y_1) = L_s(X_N - X_0) \Rightarrow 114.4(0.1364 - 0.00945) = 149.4(X_N - 0.00402)$

$$\text{i.e. } X_N = 0.1012 \quad \text{and} \quad x_N = 0.1012/(1 + 0.1012) = 0.092$$

Terminal concentrations: top, $(x_0, y_1) \rightarrow (0.004, 0.00945)$; bottom, $(x_N, y_{N+1}) \rightarrow (0.092, 0.12)$

$$\begin{aligned} \text{Operating line:} \quad G_s \left(\frac{y}{1-y} - Y_1 \right) &= L_s \left(\frac{x}{1-x} - X_0 \right) \\ \Rightarrow \quad \frac{y}{1-y} &= 1.306 \left(\frac{x}{1-x} - 0.00402 \right) + 0.00945 \end{aligned} \quad (\text{i})$$

Determination of the number of trays: The given equilibrium data and the operating line given by Eq. (i) are plotted on the x - y plane (Figure 6.12). By constructing steps between these lines (OA and BC), we get the number of ideal trays 5.5.

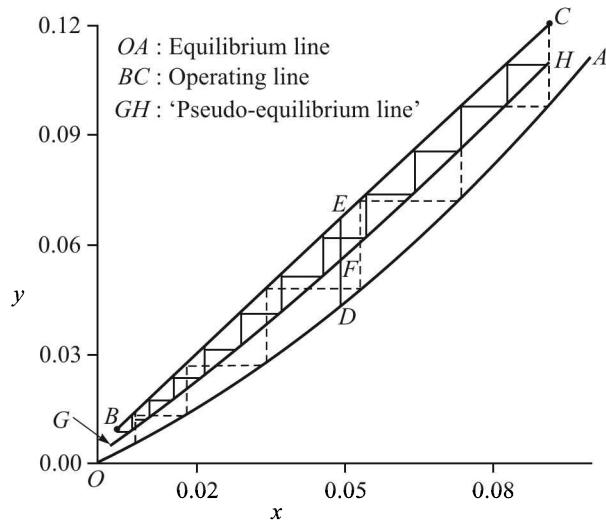


Figure 6.12 Construction of real plates, Murphree efficiency given.

To determine the *number of real trays*, we have to plot a ‘pseudo-equilibrium line’ using the given value of the Murphree efficiency. Take any vertical distance DE between OA and BC . Divide DE by the point F such that $FE = 0.45DE$. The locus of the point F is the ‘pseudo-equilibrium line’ (GH). Steps are now drawn between the operating line (BC) and the pseudo-equilibrium line (GH). The number of real stages = 12.

6.8.5 Prediction of Tray Efficiency

Prediction of tray efficiency has received a lot of attention. Besides the types of efficiency described before, a few other types were proposed by Hausen, Standart and Holland (see Lockett, 1986) although those have not been of much practical use. An excellent review of the important literature till the early 1980s has been made by Lockett. There are three methods of determination or estimation of tray efficiency: (i) totally empirical, (ii) direct scale-up from laboratory data, and (iii) semi-theoretical or theoretical models for mass transfer on a tray.

As the name suggests, the empirical method attempts to correlate the experimental tray efficiency data collected by bench scale, pilot plant, or plant data with the relevant properties and parameters of the gas–liquid systems. A chart (Figure 6.13) for ‘overall tray efficiency’ was given by O’Connell (1946) long ago and is still used. It was prepared by plotting the experimental efficiency data (collected by absorption/stripping of solutes in a number of aqueous and non-aqueous systems in columns of different diameters up to 9 ft) versus the group $mM_L\mu_L/\rho_L$. The efficiency decreases with increasing viscosity of the liquid since increasing viscosity reduces mixing and the mass transfer rate. The dependence of E_O on M_L can be explained in a similar way, but the cause of dependence on m is not very apparent. The following correlation for E_O prepared from the plot shown in Figure 6.13 is more convenient to use.

$$E_O = 1.597 - 0.199 \log(mM_L\mu_L/\rho_L) - 0.0896 [\log(mM_L\mu_L/\rho_L)]^2 \quad (6.62)$$

Here μ_L is in cP and ρ_L in lb/ft³.

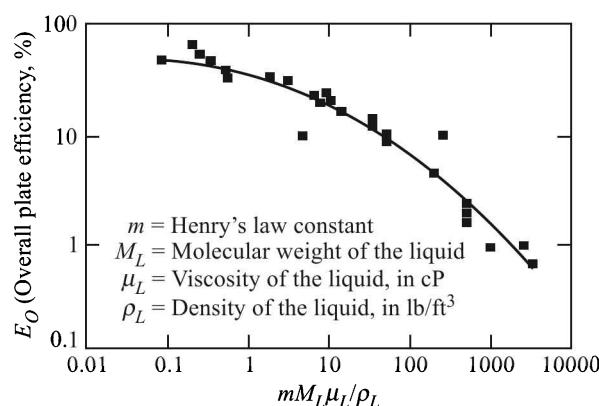


Figure 6.13 O’Connell’s chart for tray efficiency of an absorption/stripping column.

Experimental data for a system collected by using a small model column (the Oldershaw

coulmn, for example) can be scaled up to predict the efficiency in an industrial column. This technique has been used in quite a few cases.

Semi-theoretical or theoretical models for tray efficiency attempt to calculate the point efficiency E_{OG} , using equations or correlations for mass transfer coefficients or HTUs. A number of correlations for H_{IG} and H_{IL} (and also for k_G and k_L) are available in the literature (Lockett, 1986; Perry's Handbook, 7th ed., 1997, Ch. 14) may be used for this purpose. The calculated point efficiency is converted to the Murphree efficiency and then to the overall plate efficiency [$E_{OG} \rightarrow E_{MG} \rightarrow E_O$; Eq. (6.55) \rightarrow (6.57) \rightarrow (6.61)]. The assumption that the liquid is in plug flow is again an approximation. Partial 'backmixing' of the liquid on a tray, only partial mixing of the gas between adjacent trays, or existence of partially stagnant pools of liquid near the tower wall (the 'stagnant region model') occur frequently (Stichlmair and Fair, 1998) on real trays. These phenomena have to be taken into account in a model for tray efficiency. A relation between E_{OG} and E_{MG} where the gas is well mixed but the liquid is partially backmixed was reported in the AIChE Bubble Tray Design Manual of 1958. Recently Bennett et al. (1997) proposed the following correlation for the estimation of point efficiency of a sieve tray on the basis of their 'two-zone model'. They assumed a liquid-continuous zone near the tray deck, and a vapour-continuous zone above it. Mass transfer in both the zones is estimated using suitable correlations to obtain the correlation for E_{OG} .

$$E_{OG} = 1 - \exp \left[\frac{-0.0029}{1 + m \left(\frac{\rho_{mG}}{\rho_{mL}} \right) \left(\frac{D_G}{D_L} \frac{1 - \varphi_e}{A_h/A_a} \right)^{0.5}} \left[\frac{\rho_G u_h h_{Fe}}{\mu_L} \right]^{0.4136} \left(\frac{h_L}{d_H} \right)^{0.6074} \left(\frac{A_h}{A_a} \right)^{-0.3195} \right] \quad (6.63)$$

Here m = slope of the equilibrium line ($y = mx$); ρ = density; ρ_m = molar density; D = diffusivities; A_h = area of the holes; A_a = active tray area; φ_e = effective relative froth density ($= h_L/h_{Fe}$); h_L = clear liquid height on the tray; h_{Fe} = effective froth height; u_h = hole velocity of the gas; d_H = hole diameter.

The basic principles of gas absorption and design of towers have been briefly discussed in this chapter. Thermal effects play a major role in gas absorption since equilibrium relation has a strong temperature dependence. More details of absorber design, particularly design data and information like HTU and HETP values, are available in Zarzycki and Chacuk (1993), Strigle (1994), Billet (1995), and Ludwig (1997).

6.9 OTHER APPLICATIONS OF GAS ABSORPTION AND STRIPPING

Although gas absorption and stripping are common mass transfer operations in chemical process industries, there are numerous real-life situations where these phenomena play an important role. For example, absorption of oxygen in natural water is the key to survival of aquatic creatures. A blood oxygenator is a novel device (discussed in Chapter 14) for biomedical applications. Desorption or stripping of a gas has also many applications stretching beyond the boundary of process industries. Example 6.12 deals with such an application.

EXAMPLE 6.12 (VOC stripping in a shower) Exposure to volatile organic compounds has been a major human health concern in recent years. On the basis of experimental evidences, several researchers argued that important sources of exposure to VOCs occur indoor rather than the outdoor environment. One such potential source is contaminated tap water used for bathing or even washing [see Moya et al., *Environ. Sci. Technol.*, 33(1999) 1341–1349; 2321–2327. They reported experimental investigations on volatilization of contaminants in showers and dishwashers.]. A simple model of stripping and resulting accumulation of a contaminant in a shower was proposed by Little [*Environ. Sci. Technol.*, 26(1992) 1341–1349]. Let us use the following notations for the relevant parameters to make a simplified analysis of the phenomenon (see Figure 6.14): $Q_L(\text{m}^3/\text{s})$ = rate of water discharge and $V_L(\text{m}^3)$ = volume of water present ('water holdup') in the shower chamber at any time; Q_G = the rate of air exhaust through the shower; V_s = 'volume of the shower' which is taken as the volume of air in it; $K_L(\text{m}/\text{s})$ = overall desorption mass transfer coefficient; a' = total air–water contact area in the shower; C_G , C_L = concentrations (kg/m^3) of the VOC in the air and in the water in the chamber respectively at any time t ; \hat{H} = Henry's law constant for equilibrium distribution of the VOC in air and water ($C_G = \hat{H} C_L^*$). Assuming that the shower air is 'well-mixed', determine the time rate of change of VOC concentration in the shower. Also, determine the 'steady state' concentration of VOC in the shower.

Solution

We first determine the VOC concentration in the shower water as a function of time, using a 'pseudo-steady state approximation'. Since C_G is the gas-phase VOC concentration, the corresponding equilibrium water-phase concentration is $C_L^* = C_G/\hat{H}$. The overall liquid-phase driving force for desorption is $\Delta C_L = C_L - C_L^* = C_L - C_G/\hat{H}$. The following differential mass balance for the liquid-phase concentration can be written.

$$-V_L \frac{dC_L}{dt} = K_L a' (C_L - C_L^*) = K_L a' (C_L - C_G/\hat{H})$$

We integrate the above equation from $t = 0$, $C_L = C_{L,\text{in}}$ to $t = \tau$, $C_L = C_{L,\text{out}}$. Here τ is the 'average residence time' of water in the shower, $\tau = V_L/Q_L$.

$$\begin{aligned} \ln \frac{C_{L,\text{out}} - C_G/\hat{H}}{C_{L,\text{in}} - C_G/\hat{H}} &= - \frac{K_L a'}{V_L} \tau = - \frac{K_L a'}{V_L} \cdot \frac{V_L}{Q_L} = - \frac{K_L a'}{Q_L} = -\beta \text{ (dimensionless)} \\ \Rightarrow C_{L,\text{out}} &= (C_{L,\text{in}} - C_G/\hat{H}) e^{-\beta} + C_G/\hat{H} \\ \Rightarrow C_{L,\text{in}} - C_{L,\text{out}} &= C_{L,\text{in}} (1 - e^{-\beta}) - (C_G/\hat{H})(1 - e^{-\beta}) \end{aligned} \quad (\text{i})$$

Now, a differential mass balance equation for the VOC accumulation in the shower air may be

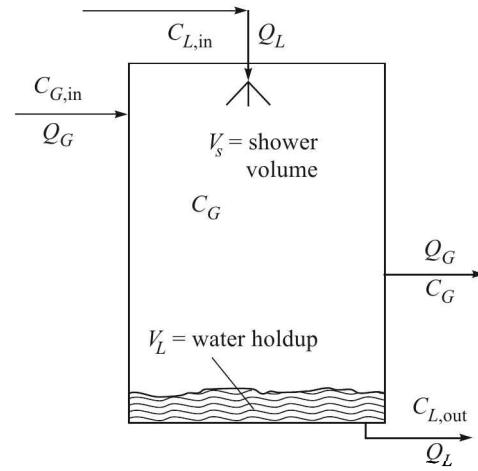


Figure 6.14 Schematic of a shower.

written taking into account the rate of desorption of VOC from water and that carried away by the exhaust air.

$$V_s \frac{dC_G}{dt} = Q_L(C_{L,in} - C_{L,out}) - Q_G(C_G - C_{G,in}); \quad C_{G,in} = \text{inlet VOC concentration in air}$$

Substitute for $C_{L,in} - C_{L,out}$ from Eq. (i) to get

$$V_s \frac{dC_G}{dt} = Q_L \left[C_{L,in}(1 - e^{-\beta}) - \frac{C_G}{\hat{H}}(1 - e^{-\beta}) \right] - Q_G(C_G - C_{G,in}) \Rightarrow \frac{dC_G}{dt} = A' - B'C_G \quad (\text{ii})$$

$$\text{where, } A' = \frac{1}{V_s} [Q_L C_{L,in}(1 - e^{-\beta}) + Q_G C_{G,in}] \quad \text{and} \quad B' = \frac{1}{V_s} \left[Q_G + \frac{Q_L}{\hat{H}}(1 - e^{-\beta}) \right]$$

Using the initial condition, $t = 0$, $C_G = C_{G,in}$, and integrating,

$$C_G = \frac{A'}{B'} - \left(\frac{A'}{B'} - C_{G,in} \right) e^{-B't} \quad \text{and} \quad C_G = A'/B' \text{ at steady state (i.e. at } t = \infty) \quad (\text{iii})$$

Equation (iii) above gives the transient VOC concentration in the shower air after the tap is turned on. It is to be noted that the above derivation based on Little's model (see the ref. given in the problem statement) is highly simplified. It does not take into account desorption from the water droplets and the pool of water separately. It considers the air to be well-mixed which does not reflect the reality. More refined models of VOC accumulation in a shower have been proposed in the literature. A two-compartment model of the shower takes care of partial mixing of the air (see Problem 6.30). Interestingly, the water 'holdup' in the shower, V_L , does not appear in Eq. (iii) above.

NOTATIONS

\bar{a}	: specific interfacial area of contact, m^2/m^3 packed volume
\bar{A}	: absorption factor, L/mG ($= L'/mG'$)
E	: fractional entrainment of liquid (see Section 6.8.3)
E_{OG}, E_{MG}, E_O	: point, Murphree, overall plate efficiency
$C, [(C)_{av}]$: solute concentration [average total molar concentration] in the liquid, kmol/m^3
$G (G_s)$: flow rate of the gas (flow rate on solute-free basis), kmol/h or kg/h
G', L'	: gas and liquid flow rates per unit area basis, $\text{kmol}/(\text{h} \cdot \text{m}^2)$
h_T	: packed height, m
H_{tG}, H_{tOG}	: height of individual and overall gas-phase mass transfer unit respectively, m
k_x, k_y	: individual mass transfer coefficients, $\text{kmol}/(\text{h})(\text{m}^2)(\Delta x)$, $\text{kmol}/(\text{h})(\text{m}^2)(\Delta y)$
k'_x, k'_y	: Colburn-Drew mass transfer coefficients (see Chapter 4)
K_G, K_L	: overall gas-phase mass transfer coefficients, $\text{kmol}/(\text{h})(\text{m}^2)(\Delta p \text{ or } \Delta C)$

K_x, K_y	: overall mass transfer coefficients
K_X, K_Y	: overall mass transfer coefficients, $\text{kmol}/(\text{h})(\text{m}^2)(\Delta X \text{ or } \Delta Y)$
$L (L_s)$: liquid flow rate (solvent flow rate, solute-free basis), kmol/h or kg/h
m, α	: Henry's law constant (in the form $y = mx$, or $Y = \alpha X$)
N_{tG}, N_{tOG}	: number of individual and overall gas-phase mass transfer unit respectively
P	: total pressure, atm, bar, mm Hg
\bar{S}	: stripping factor, $mG/L (= mG'/L') = 1/\bar{A}$
Sc	: Schmidt number
x	: solute concentration in the liquid, mol fraction
X, Y	: solute concentration in the liquid or in the gas, mol solute per mol solute-free medium
y	: solute concentration in the gas, mol fraction
y_{iBM}	: log mean concentration of the carrier (B), Eq. (6.16)
y^*	: solute concentration in the gas in equilibrium with bulk liquid of concentration x , mol fraction

SHORT AND MULTIPLE CHOICE QUESTIONS

1. The point P represents the partial pressure of a solute in the bulk gas (p_{Ab}) and the molar concentration in the bulk liquid (C_{Ab}) at a section of a column; the point M represents the interfacial composition (see Figure 6.15). The other relevant quantities and the equilibrium line are also shown. What is the slope of the line PM ?

(i) $-k_x/k_y$ (ii) $-k_L/k_G$ (iii) $-k_c/k_L$

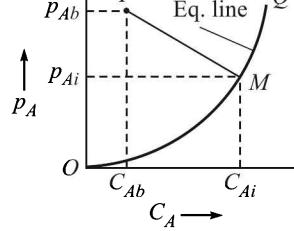


Figure 6.15

2. Equilibrium and operating lines (OP and AB) for four gas–liquid contacting operations are shown in Figure 6.16(a)–(d). (a) Identify the figures as cases of cocurrent/counter-current absorption/stripping. (b) Which of the figures represents an equilibrium co-current unit (i.e. having 100% stage efficiency)? Which one does not?

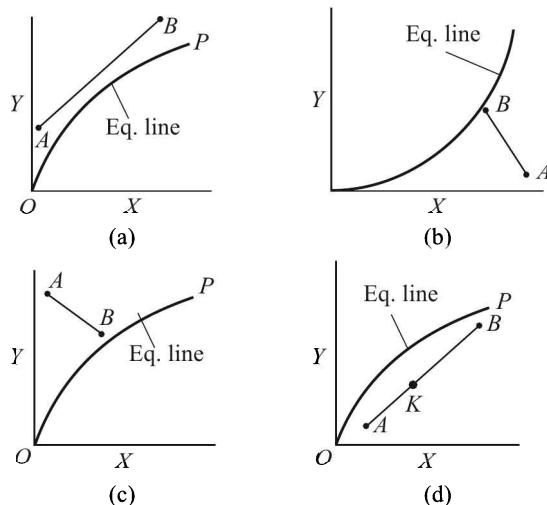


Figure 6.16 (a)–(d) A few sets of equilibrium and operating lines.

3. The equilibrium line for a gas–liquid system is shown in Figure 6.17 (admittedly it is of an unrealistic shape). The solute A is to be scrubbed in a packed tower. The top compositions (represented by the point B on the X – Y plane) are $(0, Y_2)$ and the feed gas concentration is Y_1 (in mole ratio unit, as the notation indicates). In order to determine the minimum liquid rate, a student drew a horizontal line through $Y = Y_1$ that met the equilibrium line at C . He joined BC and was wondering which one of the points C , D , and E was the pinch point. What is your answer?

(i) C (ii) D (iii) E (iv) None of these

4. The equilibrium line OP and the operating line AB in a gas–liquid contacting operation are shown in Figure 6.16(d). Indicate the overall gas-phase and liquid-phase driving forces at the point K by line segments.

5. An absorption oil loaded with benzene (mole ratio of benzene = X_1) is to be steam-stripped in a packed column. The benzene concentration in the lean oil leaving the column is X_2 . (a) If the steam rate is increased (for a fixed liquid rate and for constant values of X_1 and X_2), the terminal point P (Figure 6.18) of the operating line will shift towards the point

(i) C (ii) D (iii) E .

- (b) For the given shape of the equilibrium line, where shall the *pinch point* for minimum steam rate for stripping lie?

(i) at the top of the packed section

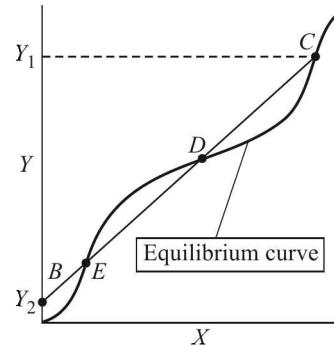


Figure 6.17

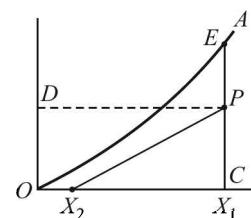
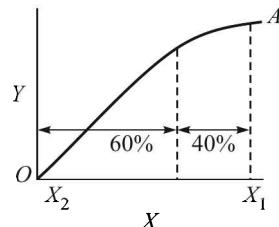


Figure 6.18

- (ii) at the bottom
 (iii) somewhere midway in the packed section.
6. A student has calculated the number of theoretical plates required for a particular gas absorption operation by using the Kremser equation. He ignored the fact that although the equilibrium line is nearly straight over the lower 60% of the concentration range involved, it is appreciably convex upwards over the upper 40% of the concentration range (Figure 6.19). In fact, he used the slope of the lower part of the equilibrium line and got N number of plates. If N' is the number of plates obtained by using a more precise method, then
- (i) $N' = N$ (ii) $N' > N$ (iii) $N > N'$.
7. A dissolved solute is to be removed from a solution by stripping with a gas in a packed tower. The height of the packing (Raschig rings) is estimated to be h under a given set of temperature conditions. Which of the following changes in the operating conditions may be done in order to reduce the packed height without reducing the degree of separation?
- (i) Increase the operating pressure
 (ii) Decrease the operating temperature and the rate of flow of the stripping gas
 (iii) Increase the temperature or decrease the pressure
 (iv) Use saddle packings rather than rings
 (v) Increase the stripping gas rate
 (vi) Add a suitable reagent to the solvent that reacts with the dissolved solute
8. How does the number of ideal trays required for *stripping* of a solute from a solution vary with the absorption factor $\bar{A} = L/mG$? With increasing \bar{A} , the number of ideal trays required to achieve a specified degree of separation will
- (i) increase (ii) decrease (iii) remain unchanged.
9. A gas is to be absorbed in a liquid of low volatility in a tray tower. Both the gas and the liquid streams are available at room temperature. The number N of real plates required has been calculated by assuming isothermal operation. But the absorption process releases considerable amount of heat and the number of ideal plates for adiabatic absorption will be greater than N . The main reason for this is:
- (i) A small amount of solvent that may vaporize will increase the gas flow rate, thereby reducing the mole fraction of the solute in the gas. This will, in turn, reduce the driving force for absorption.
 (ii) In the case of adiabatic absorption, the equilibrium line becomes steeper, and the driving force (i.e. the vertical distance between the equilibrium and the operating lines) is thus reduced.
 (iii) In adiabatic absorption, the physical properties of the phases change appreciably along the column and the number of plates required becomes more.

**Figure 6.19**

10. The adiabatic absorption referred to in Question 6.9 is proposed to be carried out in an existing unused column having N trays. The diameter of the column is also suitable to accommodate the given flow rates. The heat of absorption is not very high. Which of the following strategies seems attractive?
- (i) Cool down the solvent adequately before feeding it into the column.
 - (ii) Properly cool the feed gas stream.
 - (iii) Take out a liquid stream from one (or more) intermediate tray, cool it in a heat exchanger and feed the liquid back into the tray, thus removing the heat of absorption.
11. What is the physical significance of HTU?
- (i) It indicates inversely the ease with which a column performs separation.
 - (ii) It is the height of packing required to achieve unit change in the concentration of the solute gas.
 - (iii) It gives a measure of the interfacial area of contact in a packed column.
12. Which one of the following appears to be a representative value of the specific interfacial area of contact in an industrial scale packed tower?
- (i) 0.5 cm^{-1}
 - (ii) 0.5 ft^{-1}
 - (iii) 0.5 m^{-1}
13. What is the physical significance of NTU?
- (i) It indicates the ease of separation of a mixture.
 - (ii) It indicates the difficulty of separation of a mixture.
 - (iii) It indicates the efficiency of a particular packing for mass transfer.
14. The number of transfer units for absorption of three gases A , B and C in water are 10, 4 and 15 m respectively. The inlet and exit concentrations (mole fraction) of the gas-phase and of the liquid-phase have the same values in all the cases. For which system is the average driving force for mass transfer minimum?
15. How does the height of a transfer unit (HTU) depend on the gas and liquid flow rates (G and L)?
- (i) It increases slowly with increasing G .
 - (ii) It increases with increasing L .
 - (iii) It is independent of G but increases with L .
 - (iv) It is independent of both G and L .
- (More than one answer may be correct.)
16. A gas is being absorbed in a pure liquid and in this particular case, the liquid-side interfacial concentration is negligibly small. What fraction of the inlet gas is expected to be absorbed in one transfer unit?
- (i) 50%
 - (ii) 23.6%
 - (iii) 63.2%
17. Which of the following is a probable value of the HTU of a 3 ft diameter column packed with #25 IMTP?
- (i) 2.7 m
 - (ii) 1.3 ft
 - (iii) 105 mm

18. Which of the following quantities is appreciably dependent on the solubility of a gas in the solvent?
- (i) HTU (ii) NTU (iii) \bar{a} (specific interfacial area)
19. For absorption of a lean gas in a packed column, the absorption factor turns out to be very high. Which of the following statements is correct for this system?
- (i) The height of the overall gas-phase transfer unit is close to that of the individual gas-phase transfer units.
(ii) The height of the overall gas-phase transfer unit is very small.
(iii) The height of the gas-phase transfer unit will definitely be smaller than that of the liquid-phase transfer unit.
20. In a small section of a packed tower, the fraction of the solute absorbed corresponds to one transfer unit. The overall gas-phase driving forces (in mole fraction) at the two ends of this section are 0.014 and 0.0098 respectively. The slope of the equilibrium line is 1.1 and the absorption factor is 0.99. What is the change in the gas-phase concentration over this transfer unit?
- (i) 0.02 (ii) 0.0106 (iii) 0.012
21. Under which of the following conditions the gas-phase driving force will be constant along the column?
- (i) $L/G = 1$ (ii) $L > G$ and $m = 1$ (iii) $L/G = m$
22. In which of the following cases the number of the overall gas-phase transfer units will be equal to the number of ideal plates for the same degree of separation?
- (i) The equilibrium line and the operating line are nearly straight.
(ii) The equilibrium line and the operating line are straight and parallel.
(iii) The slope of the operating line is unity.
23. A dissolved gas is being stripped out from a dilute solution in a packed tower using steam. Concentration of the solute in the inlet liquid is 0.06 (mole fraction) and that in the lean solution at the bottom must not exceed 0.002. The equilibrium and the operating lines are nearly straight. The overall liquid-phase driving forces at the top and at the bottom of the column are found to be 0.00472 and 0.00102 respectively. What is the number of overall liquid-phase transfer units?
- (i) 24 (ii) 14 (iii) 21
24. For the absorption of a *dilute* gas in a packed tower, the heights of the individual transfer units are $H_{tG} = 1.2$ ft and $H_{tL} = 0.9$ ft. The average gas flow rate is 50 kmol/h and the water rate is 1800 kg/h. The equilibrium relation is $y = 1.5x$. What is the height of an overall liquid-phase transfer unit?
- (i) 3.5 ft (ii) 1.6 ft (iii) 0.3 ft
25. What is the physical significance of the *absorption factor A*?
- (i) It is the ratio of the slopes of the equilibrium line and the operating line.
(ii) It is ratio of the individual gas-phase and liquid-phase mass transfer coefficients.
(iii) It is fractional absorption of the feed gas.

26. A larger absorption factor means
 - (i) an increased driving force for absorption
 - (ii) an increased rate of flow of the liquid
 - (iii) an increased packed height for the same degree of separation.
27. The optimum ratio of the actual liquid rate to the minimum liquid rate for gas absorption *generally* lies between
 - (i) 0 and 1
 - (ii) 0.5 and 1.5
 - (iii) 1.2 and 2.0
28. In the design of a packed tower based on individual gas-phase and liquid-phase coefficients, the *tie lines* are found to be nearly parallel. This may occur because
 - (i) the log mean mole fraction of the carrier is nearly unity
 - (ii) the absorption is gas-film resistance controlled
 - (iii) the feed gas is concentrated.
29. In a packed column the height of an overall gas-phase transfer unit is 1.2 ft. The absorption factor is 1.6. What is the value of HETP?
 - (i) 1.2 ft
 - (ii) 1.5 ft
 - (iii) 2.8 ft
30. Which of the following is a practical range of the ‘absorption factor’ in gas absorption?
 - (i) 1 to 10
 - (ii) 0.5 to 1.5
 - (iii) 1.25 to 2.0
31. In a gas absorption column the absorption factor $\bar{A} = L/mG = 1$ and the Murphree tray efficiency is 0.60. What is the overall column efficiency?
32. A particular tray in an absorption column is equivalent to 0.8 overall gas-phase transfer units. The local absorption factor is $L/mG = 1.5$. What is the Murphree efficiency of the tray?
 - (i) 0.456
 - (ii) 0.62
 - (iii) 0.664
33. The equilibrium in a gas–liquid system follows the Henry’s law, $y = mx$. Give the range of values of the Henry’s law coefficient m , for which the equilibrium data plotted in mole ratio units [$Y = f(X)$] is
 - (i) convex downwards
 - (ii) convex upwards.
34. The equilibrium line for a gas–liquid system is given by $Y = 2.5X$. Fresh liquid enters the top of a counter-current packed bed and the gas leaves at a concentration, $Y_2 = 0.001$. The overall gas-phase driving force is $\Delta Y = 0.004$ at a section where the bulk gas concentration is $Y = 0.035$. What is the overall liquid-phase driving force (ΔX) at a section where the bulk gas concentration is $Y = 0.04$?
35. Analysis of samples drawn from the fifth tray from the top in an absorption column shows that its Murphree efficiency is 0.55. The tower diameter is not large, and the liquid-phase at a tray may be considered to be *well-mixed*. What is the number of overall gas-phase transfer units equivalent to this tray?
36. Explain how Eq. 6.24(b) can be simplified for a dilute gas–liquid system.
37. Derive Eqs. 6.24(b), (6.58), and (6.61).

38. Many random packings have multiple drip points at which the liquid film terminates to form drops. Explain qualitatively how this feature improves the mass transfer efficiency of a packing.
39. What are the physical significances of the absorption factor and the stripping factor?
40. Which of the following is a probable value of dynamic liquid holdup (volume fraction) in a packed bed?
- (i) 0.5 (ii) 0.35 (iii) 0.035
41. Which of the following is a probable value of $K_G \bar{a}$ [kmol/(h)(m³)(atm)] for absorption of NH₃ in an acidic solution in a 1 m diameter column packed with 2-inch Koch Flexiring?
- (i) 10 (ii) 200 (iii) 2000
42. If the liquid on a tray is well-mixed and the gas also gets well-mixed in the space between two consecutive trays, then
- (i) $E_{OG} = E_{MG}$ (ii) $E_{OG} > E_{MG}$ (iii) $E_{OG} < E_{MG}$
43. With increasing viscosity of the liquid, the tray efficiency generally
- (i) increases (ii) decreases (iii) remains unchanged.
44. Discuss the reasoning behind considering the quantity $G'/k_y \bar{a} (1 - y)_{iM}$ nearly constant and calling it 'height of a gas-phase transfer unit'.
45. The individual mass transfer coefficient k_y is associated with a gas-film thickness δ . How can the concept be extended to the overall coefficient K_y ? Is it possible to extend the concept to define an 'effective film thickness' corresponding to K_y in terms of the individual gas-film and liquid-film thicknesses?
46. What are the factors that influence H_{tOG} and N_{tOG} of a packed tower?
47. How can the H_{tG} and H_{tL} values of a packing available for one gas–liquid pair be modified for use in the design of another gas–liquid system?
48. Do mass transfer data collected in a wetted-wall column have relevance to mass transfer in a packed tower?
49. Comment on the assumptions made to develop Eq. (6.55) for the 'point efficiency'. How is tray efficiency affected by liquid entrainment?
50. From Eq. (6.49) obtain the following relation between the number of transfer units and the number of ideal trays N : $N_{tOG} = N \cdot (\ln \bar{S}) / (\bar{S} - 1)$. Under what condition is N_{tOG} larger (or smaller) than the number of ideal trays?
51. Which solvent/liquid would you recommend for the following gas absorption processes?
- (i) Drying of air in a H₂SO₄ plant
(ii) Separation of acrylonitrile from the reaction products of ammonoxidation of propylene
(iii) Drying of natural gas
(iv) Drying of chlorine prior to liquefaction
(v) Separation of H₂S from natural gas

- (vi) Recovery of ammonia from the 'bleed stream' from an ammonia reactor
 (vii) Separation of CO₂ from the ammonia synthesis gas
 (viii) Recovery of light hydrocarbons from a refinery vent stream.
52. Is there a value of the absorption factor, $\bar{A} = L/mG'$, for which the Murphree efficiency (for the liquid in plug flow) is the same as the point efficiency?

PROBLEMS

- 6.1 (*Minimum liquid rate for absorption*)¹ A packed tower is to be designed for the absorption of 98% of the ammonia(A) from an air–ammonia mixture containing 4% ammonia at a rate of 4200 Nm³ (normal cubic metre) per hour using water as the solvent. The tower operates at 105.1 kPa and 303 K. The equilibrium data for NH₃–water system at 303 K [Sherwood, T.K., *Ind. Eng. Chem.*, 17(1925) 745] are given below:

Partial pressure of NH ₃ (mm Hg)	19.3	29.6	40.1	51.0	79.5	110
kg NH ₃ per 100 kg water	2	3	4	5	7.5	10

- (a) Calculate and plot the equilibrium data as x_A vs. p_A , x_A vs. y_A and X_A vs. Y_A . Up to what value of p_{NH_3} can the Henry's law be used to describe the equilibrium?
 (b) Calculate the minimum liquid rate for the absorption (the inlet water is NH₃-free).

- 6.2 (*Flooding velocity; number of trays*)² A stream of nitrogen gas containing 9.5% benzene by volume is to be scrubbed in a packed tower with an absorption oil in order to reduce the benzene content to 0.5% by volume before the gas can be recycled. The feed gas rate is 1550 Nm³/h; it enters the column at 308 K. The column operates at 308 K and 1.07 bar pressure. Benzene-free oil is fed to the column at a rate of 4600 kg/h. The equilibrium distribution of benzene between the gas and the oil is given below.

$$Y = \frac{0.19 X}{1 + 0.81 X}$$

where X = moles of benzene per mole of benzene-free oil, and Y = moles of benzene per mole of nitrogen. The properties of the oil are: density = 840 kg/m³, viscosity = 4.5 cP, and mol. wt. = 230.

- (a) Do you think that the system is ideal (i.e. obeys the Raoult's law) in respect of equilibrium? Justify your answer by calculations. Can you calculate the vapour pressure of benzene at the given temperature?
 (b) Calculate the mole fraction of benzene in the oil at the bottom of the tower.
 (c) Calculate the diameter of the tower if the superficial gas velocity is 0.32 m/s. If the tower is packed with 1½ inch ceramic Raschig ring, calculate the ratio of the actual gas rate to that at flooding.
 (d) If the separation is done in a plate column and the overall tray efficiency is 35%, determine the number of real trays required. [Note: Tray efficiency for absorption is often found to be substantially lower than for distillation.]

- 6.3 (Number of trays in a stripping tower)²** The absorption oil loaded with CS₂ leaving the tower described in Example 6.6 is to be subjected to steam stripping in a tray tower. The pressure in the column is 1.12 bar and the temperature is 105°C. The steam is essentially saturated. Vapour pressure of CS₂ at this temperature is 4.7373 bar. Calculate the number of ideal trays required if the stripped oil leaving the bottom of the tower does not have more than 50 ppm of CS₂. The steam rate used is 1.3 times the minimum. The Raoult's law is applicable. Note that the bottom liquid has a very low concentration of the solute and graphical construction may be difficult in this low concentration region.
- 6.4 (Calculation of N_{log} and packed height)²** Sulphur dioxide is to be scrubbed from an air stream in a small packed tower by contacting it with an organic amine. The feed gas 3% SO₂ by volume, and 95% of it is to be absorbed. The total gas rate is 150 m³/h at 20°C and 1.1 bar absolute pressure. The liquid enters the column at a rate of 1.40 kmol/h. Given: the overall mass transfer coefficient, $K_G = 3.2 \times 10^{-4}$ kmol/(m²)(s)(Δp, bar); the effective gas–liquid contact area = 105 m² per m³ of packed volume; slope of the equilibrium line, $m = 0.17$. Determine the number of overall gas-phase mass transfer units and the packed height if the column is 1 ft in diameter.
- 6.5 (Use of Kremser equation)²** Groundwater contaminated with 1,1,1-trichloroethane (TCA)[#] is to be air-stripped at 25°C and 1.1 bar pressure at a rate of 9000 kg/h. The feed water has 600 ppm of the VOC. A stripping column of suitable diameter with eight trays having 40% overall efficiency is available. The exit stripping air must not contain more than 2.5% TCA by volume. The equilibrium distribution of TCA between water and air has been reported by Turner et. al. [AIChE J, 42(1996) 1772–1788].

$$K' = 0.204 + 0.0182\theta + 0.000173\theta^2 \frac{\text{mg/lit, air}}{\text{mg/lit, water}}; \quad \theta \text{ in } ^\circ\text{C}.$$

Calculate (a) the Henry's law constant (i.e. ' m ' in $y = mx$) for TCA at 25°C and (b) the percentage removal of TCA achievable in the column.

- 6.6 (Use of Colburn-Drew coefficients)²** A packed column receives 10 kmol/h·m² of feed gas (8 mol% solute, 92% carrier) and 30 kmol/h·m² of fresh solvent. It is required to remove 97% of the solute. The equilibrium relation for the gas–liquid system is $y = 2.5x + 4.5x^2$, $x \leq 0.08$. Is the solvent rate more than the minimum? Calculate the value of the absorption factor at a section of the column where the bulk liquid concentration is $x = 0.02$? If the individual 'Colburn-Drew mass transfer coefficients' are $k'_x = 1.2$ kmol/(m²)(s)(Δx) and $k'_y = 0.15$ kmol/(m²)(s)(Δy), calculate the interfacial concentrations and the individual gas-phase driving force.
- 6.7 (Effect of column pressure on the extent of separation)³** An absorption column is to be designed for reducing the concentration of a toxic vapour in an air emission from 1% to 0.02 mole%. The column will operate at 20°C and 0.5 bar gauge pressure. The scrubbing liquor flow rate is 1.3 times the minimum. The gas–liquid equilibrium relation ($p = Hx$) is linear and the Henry's law constant is 10 bar. (a) How many trays are required

[#]Chlorinated hydrocarbons such as trichloroethane and trichloroethylene were used in large quantities in the past for cleaning of metal parts and sometimes for dry cleaning of garments. Spills and leakages of such compounds have been a cause of contamination of groundwater at some places of a few countries like the US.

to achieve this separation if the overall tray efficiency is 40%? (b) After the column is designed, it becomes necessary to reduce the concentration of the toxic vapour in the vent air to 0.01% in order to meet a modified pollution control regulation. Is it possible to achieve this goal by operating the column at the same temperature but at an enhanced pressure? If so, at what pressure should the column be operated?

- 6.8 (Packed tower calculations)²** Formaldehyde present in an effluent air stream is to be absorbed into an effectively non-volatile solvent at 1 atm total pressure and 30°C. The inlet and outlet concentrations of the organic in the air stream are 4.25% and 0.08 mole% respectively. The solvent entering the column has only traces of formaldehyde. The ratio of molar rates of liquid and gas inputs is 2.9 and the equilibrium relation is $y = 2.5x$. Determine: (a) the equation of the operating line, (b) the number of overall gas-phase transfer units, and (c) the maximum admissible ratio of the gas and liquid rates for this separation.
- 6.9 (Use of Murphree efficiency)³** Benzene from a stream of N₂ is to be separated in an absorption column followed by steam stripping to recover the hydrocarbon. The feed (100 kmol/h) enters the absorption column at 25°C and is 90% saturated with benzene. Not more than 2.5% of the feed benzene should leave the column. The feed solvent has 0.007 mole benzene per mole of 'pure' solvent (mol. wt. = 240). The loaded solvent leaving the absorber has 0.3 mole benzene per mole of pure solvent. The solution is heated and fed to the stripping tower to reduce the benzene content to 0.007 mole ratio so that the lean liquid can be cooled and recycled to the absorption. The steam rate used is 1.5 times the minimum. Both the towers operate at atmospheric pressure. The absorption tower operates at 25°C and the stripping tower at 110°C. The equilibrium data at 25°C are given below.

X	0.04	0.08	0.12	0.16	0.20	0.24
Y	0.011	0.0215	0.032	0.040	0.051	0.060

The Raoult's law may be used to determine the equilibrium relationship at the stripping temperature of 110°C. Vapour pressure of benzene can be calculated from the Antoine equation,

$$\ln P^v = 17.5818 - (3867/T); \quad P^v = \text{vapour pressure in mm Hg}; \quad T = \text{temperature, K}.$$

(a) Calculate the liquid rate to the absorber. (b) Determine the number of ideal trays required for absorption. (c) What is the minimum possible steam rate? (d) Determine the number of ideal trays in the stripping column. (e) How many real trays are required for stripping if the Murphree plate efficiency is 50%?

- 6.10 (Effect of pressure; packing size)³** A gaseous effluent stream containing 2 mole% ammonia is scrubbed with water at 30°C and 1.2 bar total pressure to remove 99% of the solute in a tower packed with 1-inch intalox saddle. The feed water may have 0.00008 mole fraction of ammonia in it. The gas rate is 1200 normal cubic metre (Nm³) per hour and the equilibrium relation is $p = x (p \text{ in bar})$. (a) If 52 kmol/h·m² water is used, calculate the concentration of NH₃ in the water leaving the tower in kmol/m³ (the density of the liquid is the same as that of water). (b) What is the Henry's law constant for the given form of the equilibrium line? (c) Determine the mass transfer driving force

- (gas-phase basis) at a section of the column where 50% of the inlet gas has been absorbed. (d) What will be the effect of: (i) doubling the operating pressure, and (ii) using the $\frac{1}{2}$ -inch size saddles as packing on the fractional removal of the solute?
- 6.11 (Calculation of liquid rate in an existing tower)²** A solute is removed from a gas stream by using an existing absorption tower filled to 8.6 m by IMTP. The tower diameter is considered suitable for the required service, but the water rate has to be calculated. The feed gas has 3.5 mole% of the solute and the exit gas may have 0.05 mole% of it. The solubility of the solute is given by Henry's law: $y = 2.53x$. Following data are available: superficial gas velocity at the top of the column = 0.8 m/s; total pressure = 1 atm; temperature = 26°C; overall gas-phase mass transfer coefficient, $K_G \bar{a} = 7050 \text{ kg}/(\text{h})(\text{m}^3)(\Delta p, \text{ bar})$; molecular weight of the solute = 30. Determine the liquid flow rate and HETP of the packing.
- 6.12 (A simple design problem)³** It is required to design a packed tower to treat 40,000 ft³/h of an air stream containing 20 mole% of SO₂ at 70°C and 1 atm total pressure. It is necessary to recover 96% of the SO₂ using water at a rate 30% more than the minimum. The column may be packed with 1½-inch Raschig rings and may be operated at 60% of the flooding velocity. The individual mass transfer coefficients are: $k_G a = 0.028(G)^{0.7}(L')^{0.25}$ and $k_L a = 0.044(L')^{0.82}$ where G' and L' are the gas and liquid mass flow rates, in lb/h·ft². Design the tower.
- 6.13 (A simple case of solvent selection)³** It is required to absorb 97% of the solute from a feed gas of concentration 8% by volume. The gas enters the column at a rate of 100 kmol/h. A column of suitable diameter with ten ideal trays is available. It is necessary to select one of two solvents otherwise suitable. The following data are supplied. Solvent 1: molecular weight = 200; equilibrium relation, $Y = 2.5X$; cost, Rs 50 per kg. Solvent 2: molecular weight = 130; equilibrium relation, $Y = 1.5X$; cost, Rs 30 per kg. Pumping and spillage loss is estimated to be 0.001% of the circulation rate for both the solvents. But solvent 2, though cheaper, is volatile and 0.004% of it is lost in the stripping column. Which solvent will you recommend considering the above costs only?
- 6.14 (Packed tower calculations)²** Hydrogen sulphide has to be removed from a light refinery hydrocarbon stream before the gas is subjected to further processing. The feed gas contains 0.035 mole fraction of H₂S, 99% of which has to be removed by scrubbing with an aqueous weakly basic solution. The 1.5 m diameter absorption column operates at 25°C and 101.3 kPa total pressure. The feed gas rate is 45 kmol/h·m² and the concentration of the solute(H₂S) in the absorbent solution leaving the tower is 0.015 mole H₂S per mole of the H₂S-free liquid. The equilibrium relation is linear and is given by $y = 1.95x$ and the overall gas-phase film coefficient is $K_y \bar{a} = 130 \text{ kmol}/(\text{m}^3)(\text{h})(\Delta y)$. Calculate (a) the rate of solvent flow, (b) the slope of the operating line in small concentration range if mole fraction units are used in the calculations, (c) the overall driving forces at the top and at the bottom of the tower on (i) gas phase basis and (ii) liquid phase basis, (d) the number of transfer units, and (e) the packed height.
- 6.15 (Design of a column receiving two feeds)³** A tray tower is to be designed for processing two feed streams having the same solute and carrier gases but at different concentrations. Feed 1 enters at a rate of 90 kmol/h with 10 mol% of the solute.

Feed 2 has a flow rate of 60 kmol/h with 5 mol% solute. The equilibrium relation is linear, $Y = 1.2X$. It is required to absorb 95% of the total solute. What is the minimum liquid rate for the separation? If 1.25 times the minimum liquid rate is used, determine the number of ideal trays.

- 6.16 (Packed tower calculations)²** A binary gas mixture containing 7.5% of a solute *A* has to be scrubbed with the solvent *B* in a packed tower. Based on the flooding calculations, a tower diameter of 1.2 m has been selected. The total gas rate at the bottom is 60 kmol/h. The exit gas must not contain more than 0.2% of the solute. Solute-free liquid entering the column at the rate of 40 kmol/h is used. The gas-phase and the liquid-phase mass transfer coefficients (based on the mole ratio unit of concentration) are $k_X = 2.05 \text{ kmol}/(\text{m}^2)(\text{h})(\Delta X)$, and $k_Y = 1.75 \text{ kmol}/(\text{m}^2)(\text{h})(\Delta Y)$. The equilibrium line is linear, $Y = 0.63X$. The specific interfacial area of gas–liquid contact is $71 \text{ m}^2/\text{m}^3$. Calculate the height of packing necessary for the separation. By what per cent must the packed height be increased in order to remove 99.5% of the solute? Determine the slope of the operating line in each case.
- 6.17 (A column having trays above the packing)³** A mixture containing 10% of the solute *A* is to be scrubbed with a solvent in order to remove 98% of it. The absorption column has three ideal trays at the top below which there is a packed section. The gas flow rate at the bottom is $70 \text{ kmol}/(\text{m}^2 \cdot \text{h})$, and the liquid leaving the column has 8.2 mol% solute in it. The equilibrium relation is given by $y = 1.05x + 0.32x^2$, $x \leq 0.1$. If the overall volumetric mass transfer coefficient is $K_y \bar{a} = 95 \text{ kmol}/(\text{h})(\text{m}^3)(\Delta y)$, calculate the height of the packed section.
- 6.18 (Calculation of packed height and HETP)³** Benzene is absorbed from a stream of air into a non-volatile oil in a countercurrent packed tower at 299 K and 105 kPa. The equilibrium is approximately linear, $y = 0.19x$. The solvent (mol. wt. = 240) enters the column at a rate of $8500 \text{ kg}/(\text{h} \cdot \text{m}^2)$. Terminal concentrations of benzene in the liquid and the gas phases are: $x = 0.0008$, $y = 0.0005$ at the top, and $x = 0.152$, $y = 0.038$ at the bottom. The heights of individual gas and liquid phase transfer units are $H_{IG} = 0.33 \text{ m}$ and $H_{IL} = 0.43 \text{ m}$. Calculate (a) the heights of overall gas-phase and the overall liquid-phase transfer units, and (b) the HETP of the packing. Equations 6.24(c) and (6.49) may be used.
- 6.19 (Use of Billet's correlations for tower design)³** A stream of air having 3% acetone vapour is to be scrubbed with water at a rate of $6000 \text{ m}^3/\text{h}$ at 27°C and essentially atmospheric pressure in a packed tower of 1 m^2 cross-section in order to remove 98% of acetone. The tower should use 2-inch metal Pall ring packing. Water is supplied at a rate 1.3 times the minimum. A linear equilibrium relation, $y = 2.31x$, is applicable. The mass transfer coefficients (k_L and k_c) can be calculated using the Billet's correlations, Eqs. (6.39) and (6.40). The parameters of the packing are given in Tables 5.9 and 6.4. Following physical property data may be used. Gas: $\mu_G = 1.813 \times 10^{-5} \text{ kg}/(\text{m.s})$; $D_G = 1.08 \times 10^{-5} \text{ m}^2/\text{s}$; liquid: $\mu_L = 8.57 \times 10^{-4} \text{ kg}/(\text{m.s})$; $D_L = 1.18 \times 10^{-9} \text{ m}^2/\text{s}$; $\rho_L = 997 \text{ kg}/\text{m}^3$; $\sigma_L = 0.072 \text{ kg}/(\text{s}^2)$. Calculate the required height of packing.
- 6.20 (How N_{IG} increases with the degree of removal of the solute)²** A stream of air containing 5% by volume of ammonia is to be scrubbed with dilute H_2SO_4 in a packed

tower. The gas and the liquid rates are the same, $1200 \text{ lb/h}\cdot\text{ft}^2$. The overall gas-phase mass transfer coefficient is $18 \text{ lbmol}/(\text{ft}^3)(\text{h})(\text{atm})$. Prepare a plot of N_{tOG} versus the fractional removal of ammonia from the gas (starting with 93% removal).

- 6.21 (Calculations for a particular tray)²** Following data on concentration of the gas and the liquid streams entering the p th tray of a gas absorption column are available: $y_{p+1} = 0.08$, $x_{p-1} = 0.044$. The equilibrium relation is $y^* = 1.8x^{1.2}$. The ratio of liquid and gas flow rates on solute free basis (L_s/G_s) in the column is 1.3. If the p th tray has a Murphree efficiency of 0.55 [expressed in terms of mole fraction unit, see Eq. (6.56)], calculate the concentrations of the streams leaving the tray.
- 6.22 (Calculation of point efficiency)²** The gas flow rate to the p th tray of an absorption tower is $120 \text{ kmol/h}\cdot\text{m}^2$. The froth depth is 0.28 m and it is equivalent to 0.4 transfer units. Calculate the overall volumetric gas-phase mass transfer coefficient and the point efficiency. The equilibrium relation is given by $y = 2.3x$.
- 6.23 (Tray tower calculations)³** A solute is to be steam-stripped from a solution containing 0.08 mole fraction of A in order to remove 98% of the solute. At the temperature and pressure of the column the equilibrium relation is $Y = 2.08X$. (a) Determine the terminal points of the operating line on the $X-Y$ line for minimum steam rate. (b) If 1.3 times the minimum liquid rate is used, determine the equation of the operating line and the number of ideal trays required for removal of 98% of the solute. (c) Calculate the change in concentration of the stripping gas (steam) in mole fraction as it passes through the fifth tray from the top. Calculate algebraically; do not use a graph paper.
- 6.24 (Calculation of HETP)²** A packed absorption tower operates at an average absorption factor of 1.35. The liquid flow rate is $65 \text{ kmol/h}\cdot\text{m}^2$ and the equilibrium relation is $y = 1.5x$. The overall gas-phase mass transfer coefficient is $K_G \bar{a} = 0.0113 \text{ kmol}/(\text{s})(\text{m}^3)$ (Δp , bar) and the total pressure is 1.05 bar. Calculate the HETP of the packing if the feed gas is 'dilute'.
- 6.25 (Stripping of VOC)³** A column provided with a 3-m packed bed is being used to air strip trichloroethylene (TCE) in contaminated water (400 ppm) in order to reduce the concentration to 3 ppm. The air rate is twice the minimum. Is it possible to strip TCE from contaminated water having 600 ppm of the VOC down to 0.1 ppm in the same column with an enhanced packed height, keeping the water and air rates unchanged? What packed height do you suggest? The equilibrium relation is $y = 661.1x$ (Yaws et al., *Chem. Eng. Progr.*, Feb 2005, 50–56).
- 6.26 (Efficiency calculation)³** Following data are available for a particular tray in a gas absorption column: $G' = 160 \text{ kmol/m}^2\cdot\text{h}$; $k_G = 4.7 \text{ kmol}/(\text{m}^2)(\text{h})(\text{atm})$; $k_L = 5.56 \text{ m/h}$; total pressure, $P = 1.3 \text{ atm}$; froth height, $z_L = 0.2 \text{ m}$; absorption factor, $\bar{A} = 1.2$; specific interfacial area, $\bar{a} = 100 \text{ m}^2/\text{m}^3$; and the equilibrium relation, $p = H'C$ [$H' = 0.3 \text{ atm}/(\text{kmol/m}^3)$]. Calculate (a) the point efficiency; (b) the Murphree efficiency if the liquid is in plug flow over the tray with (i) no entrainment, and (ii) for an entrainment rate of $E = 0.06$; (c) the overall column efficiency if the liquid is in plug flow on the tray and there is no entrainment.

- 6.27 (Number of ideal stages given; calculate the liquid rate)³** A feed having 12 mol% solute is to be scrubbed at a rate of 8000 m³/h (27°C; 1 atm) with a ‘pure’ solvent. The targeted removal of the solute is 96%. An existing tower with an equivalent of ten ideal trays is available for the purpose. Calculate the liquid rate required as well as the exit liquid concentration. The equilibrium relation is $y = 0.5x$.
- 6.28 (Number of trays given; side stream withdrawal)³** Three hundred kilomole per hour of a gas mixture is to be scrubbed in an existing column. The feed contains 10 mol% of a solute, 95% of which is to be absorbed. A side stream of the liquid having 10 mol% of the solute is to be withdrawn. The column has a total of ten ideal trays. There is a nozzle at the sixth ideal tray from the top which is to be used for side-stream withdrawal. The equilibrium relation is $y^* = 0.5x$. Determine the required liquid rate to the column, and the total moles of side stream that may be withdrawn per hour.
- 6.29 (A case of side stream withdrawal)³** An absorption column receives 225 kmol of feed gas having 10 mole% of a solute and it is desired to remove 98% of it. A solvent rate of 280 kmol/h is suggested. The feed solvent is solute-free. A side stream having 4 mol% of the solute is to be withdrawn from an appropriate plate at a rate of 75 kmol/h for use in another processing unit in the plant. Determine the total number of ideal trays required and the tray from which the side stream should be withdrawn. The equilibrium relation for the system is $Y = 1.1X$.
- 6.30 (A two-compartment model of a shower)³** To develop a seemingly more realistic model of accumulation of a contaminant in a shower, assume that the shower stall is separated from the rest of the bathroom by a partition that ends a little below the ceiling of the room. While the net rate of exhaust of air from the bathroom remains at Q_G m³/s, an exchange of air between the two compartments occurs at a rate of Q_{Ge} m³/s. The arrangement is schematically shown in Figure 6.20.

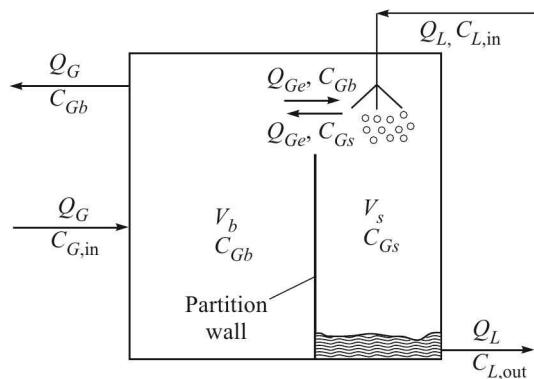


Figure 6.20 Schematic of the two-compartment model.

Develop an expression for the transient VOC concentration in the shower based on the above model. Following data are given: $V_s = 2.8 \text{ m}^3$; $V_b = 8.1 \text{ m}^3$; $Q_L = 0.0137 \text{ m}^3/\text{min}$; $Q_{Ge} = 0.11 \text{ m}^3/\text{min}$; $Q_G = 0.0378 \text{ m}^3/\text{min}$; $K_{La'} = 0.029 \text{ m}^3/\text{min}$; $C_{L,in} = 3.4 \mu\text{g/litre}$; $C_{G,in} = 0.0014 \mu\text{g/litre}$; equilibrium relation, $C_G = \hat{H}C_L^*$; $\hat{H} = 0.57$.

- 6.31 (A cell model for the point efficiency)³** In the derivation of Eq. (6.55) for the point efficiency, it has been tacitly assumed that the liquid is *vertically well-mixed* at any location on a tray but the gas is in *plug flow*. However, expectedly there will be some backmixing in the gas phase. So it may be a more realistic approach to consider that the depth of the dispersion (h) is divided into j -number of cells and the gas is *well-mixed* in each of the cells. In effect, the gas concentration changes in steps, not continuously. The liquid is considered vertically well-mixed as before. Show that the modified model leads to the following expression for the point efficiency [Kunesh et al., *Ind. Eng. Chem. Res.*, 35(1996) 2660–2671].

$$E_{OG} = 1 - \left(1 + \frac{N_{tOG}}{j}\right)^j$$

It may be noted that if the number of cells is large ($j \rightarrow \infty$), the gas phase is virtually in plug flow and the above equation reduces to Eq. (6.55).

- 6.32 (Murphree liquid tray efficiency and the number of real trays)³** For liquid-phase mass transfer controlled systems, the Murphree liquid tray efficiency is sometimes preferred. For a stripping column, we define it as

$$E_{ML} = \frac{x_n^* - x_{n+1}}{x_n^* - x_n}; \quad y_n = mx_n^*$$

If we assume that: (i) the gas and the liquid flow rates do not change substantially, (ii) $\bar{A} = L/mG$ is small, (iii) the stripping gas is solute-free $y_{N+1} = 0$, and (iv) the concentration of the solute in the lean solution is small ($x_n \ll x_0$), show that the number of ‘real trays’ required for the separation is given by (see Kunesh et al., in Problem 6.31)

$$N = \frac{\ln(x_N/x_0)}{\ln(1 - E_{ML})}$$

- 6.33 (Absorption of a dilute gas in an existing tower)²** A waste gas stream containing 2 mole% ammonia is to be scrubbed in a packed tower using dilute sulphuric acid. The gas rate is 92 kmol/h and 2.2 mole of the scrubbing liquid is to be used per mole of the gas. A packed tower of 1 m diameter having 3 m bed of a 25 mm size proprietary plastic packing is available. The height of an overall gas phase transfer unit is given as $H_{tOG} = 0.0484(G')^{0.15}(L')^{-0.2}$ [H_{tOG} in metre; G' and L' in kmol/m²·h]. It is required to reduce the ammonia concentration to 10 ppm. Will the tower be suitable for the purpose? An average gas rate and the feed liquid rate can be used for H_{tOG} calculation.

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7

Distillation

Distillation is the technique of preferential separation of the more volatile component(s) from the less volatile ones in a feed solution by partial vaporization of the feed followed by condensation. The vapor produced is richer in the more volatile component(s). The distribution of the components in the two phases is governed by the vapour–liquid equilibrium relationship. Only partial separation of the components can be achieved in this way. In order to have a larger degree of separation, multistage contact between the vapour and the liquid phases is arranged in a suitable device called a ‘distillation column’ (for example, a tray column or tower described in Chapters 1 and 5). The most important parts of a distillation unit as illustrated in Figure 7.1 are: (a) a column having trays or a packing and suitable internals (discussed in Chapter 5), (b) a reboiler, and (c) a condenser. The feed enters at a suitable point of the tower. The reboiler, normally heated by steam, partially vaporizes the liquid received from the bottom of the column. The vapour flows up through the trays or through the packing in the column, leaves at the top and enters into an overhead condenser. A part of the condensate is withdrawn as the *top product* and the rest is fed back into the column as *reflux* which flows down the trays or the packing. An intimate contact between the downflowing liquid and the upflowing vapour occurs on the trays or on the packing surface. Exchange of mass takes place between the liquid and the vapour phases. The more volatile components move from the liquid to the vapour phase and the less volatiles move in the reverse direction—from the vapour to the liquid phase. As a result, the concentrations of the more volatiles gradually increase in the vapour as it goes up, and the concentrations of the less volatiles increase in the liquid phase as it flows down the column. In this way, a higher degree of separation of the more volatiles from the less volatiles is achieved. The top product drawn from the condenser is rich in the more volatiles. The bottom product has a high concentration of the less volatiles and only a small amount of the more volatiles. The more the number of trays (or the depth of packing) and the more the *reflux ratio* (it is defined in Figure 7.1), the better is the separation. A distillation column described above performs the job of ‘fractional distillation’, i.e. separation of a feed into two or more products differing in boiling points or boiling ranges. Distillation is the most mature, albeit energy-intensive, technique of separation of liquid mixtures in small to large capacities. Columns as big as 10 m in diameter and having 150 or more trays have been reported to be in operation.