

Chapter 10

Absorption and Stripping

INTRODUCTION

The removal of one or more selected components from a gas mixture by absorption is an important operation in engineering. The process of absorption conventionally refers to the intimate contacting of a mixture of gases with a liquid so that part of one or more of the constituents of the gas will dissolve in the liquid. The contact usually takes place in some type of packed or plate column. This chapter will therefore deal exclusively with packed or plate equipment. Only equipment and design procedures are emphasized, as a detailed presentation of the theory, including diffusional process, mass transfer coefficients, equilibrium (lines), operating lines, etc., has already been covered in Chapters 6–8.

Since gas absorption is concerned with the removal of one or more species from a gas stream by treatment with a liquid, necessary information includes the solubility of these constituents in the absorbing liquid. In gas absorption operations, the equilibrium of interest is that between a nonvolatile absorbing liquid (*solvent*) and a *solute* gas. The solute is ordinarily removed from its mixture in a relatively large amount of a carrier gas that does not dissolve in the absorbing liquid.

Temperature, pressure, and the concentration of solute in one phase are independently variable. The equilibrium relationship of importance, again, is a plot of x , the mole fraction of solute in the liquid, against y (or y^*), the mole fraction in the vapor in equilibrium with x . Thus, for cases which follow Henry's law (see Chapter 6), Henry's law constant, m , can be defined by

$$y^* = mx \quad (10.1)$$

or the equivalent $y = mx^*$.

The engineering design of gas absorption equipment must be based on a sound application of the principles of diffusion, equilibrium, and mass transfer. The main requirement in equipment design is to bring the gas into intimate contact with the liquid, i.e., to provide a large interfacial area and a high intensity of interface renewal, and to minimize resistance and maximize driving force. This contacting of the phases can be achieved in many different types of equipment, the most important of which are

either packed or plate columns. The final choice often rests with the various criteria that may have to be met. For example, if the pressure drop through the column is large enough such that horsepower costs become significant, a packed column may be preferable to a plate-type column because of the lower pressure drop. Again, primary emphasis in this section is placed on packed and plate columns.

In most processes involving the absorption of gaseous constituents from a gas stream, the gas stream is the process fluid; hence, its inlet conditions (flow rate, composition, and temperature) are usually known. The temperature and composition of the inlet liquid and the composition of the outlet gas are also usually specified. The main objectives, then, in the design of an absorption column, are the determination of the solvent (liquid) flow rate and the calculation of the principal dimensions of the equipment (column diameter and height). These three topics are reviewed sequentially later in this chapter.

The general design procedure consists of a number of steps that have to be taken into consideration (details of which follow shortly). These include^(1,2):

- 1 Solvent selection
- 2 Equilibrium data evaluation
- 3 Estimation of operating data (usually obtained from a mass and energy balance, where the energy balance determines whether the absorption process can be considered isothermal or adiabatic)
- 4 Column selection (should the column selection not be obvious or specified, calculations must be carried out for the different types of columns, and the final selection based on economic considerations)
- 5 Calculation of column diameter (for packed columns this is usually based on flooding conditions, and for plate columns is based on the optimum gas velocity or the liquid handling capacity of the plate)
- 6 Estimation of the column height or the number of plates (for a packed column, the column height is obtained by multiplying the number of transfer units, obtained from a knowledge of equilibrium and operating data by the height of a transfer unit; for plate columns, the number of theoretical plates, often determined from the plot of equilibrium and operating lines, is divided by the estimated overall efficiency to give the number of actual plates, which in turn allows the column height to be estimated from the plate spacing)
- 7 Determination of pressure drop through the column (for packed columns, correlations dependent on packing type, column operating data, and physical properties of the constituents involved need to be available to estimate the pressure drop through the packing; for plate columns, the pressure drop per plate is obtained and multiplied by the number of plates)

Although detailed absorber calculations will be provided, a brief introduction follows. The usual operating data to be provided or estimated are the flow rates, terminal concentration, and terminal temperature of the phases. The flow rates and concentrations fix the operating line, while the terminal temperatures provide an indication as to what extent the operation can be considered isothermal (i.e., whether the

equilibrium line needs to be corrected for changes in liquid temperature). The operating line is obtained by a mass balance, and the outlet liquid temperature is evaluated from an energy balance on the column. In applications where relatively small quantities of gaseous constituents are being absorbed, temperature effects are usually negligible.

In gas absorption operations, the choice of a particular solvent is also important. Frequently, water is used as it is inexpensive and plentiful, but the following properties must also be considered.

- 1 Gas solubility—a high gas solubility is desired since this increases the absorption rate and minimizes the quantity of solvent necessary; generally, a solvent of a chemical nature similar to that of the solute to be absorbed will provide good solubility
- 2 Volatility—a low solvent vapor pressure is desired since the gas leaving an absorption unit is ordinarily saturated with the solvent and much will therefore be lost
- 3 Corrosiveness
- 4 Cost (particularly for solvents other than water)
- 5 Viscosity—low viscosity is preferred for reasons of rapid absorption rates, improved flooding characteristics, lower pressure drops, and good heat transfer characteristics
- 6 Chemical stability—the solvent should be chemically stable and, if possible, nonflammable
- 7 Toxicity
- 8 Low freezing point—if possible, a low freezing point is favored since any solidification of the solvent in the column could prove disastrous

Once the solvent is specified, the choice (and design) of the absorption system may be determined.

DESCRIPTION OF EQUIPMENT

The principal types of gas absorption equipment may be classified as follows:

- 1 Packed columns (continuous operation)
- 2 Plate columns (stage operation)
- 3 Miscellaneous

Of the three categories, the packed column is most commonly used. Note that plate columns received treatment in the previous chapter (Distillation).

Packed Columns

Packed columns are usually vertical columns that have been filled with packing or material of large surface area. The liquid is distributed over and trickles down through the packed bed, thus exposing a large surface area to contact the gas. The

countercurrent packed column (see Fig. 10.1) is the most common unit encountered in gaseous removal or recovery. A photograph of this unit in the Unit Operations Laboratory at Manhattan College is provided in Figure 10.2.

The gas stream moves upward through the packed bed against an absorbing or reacting liquor (solvent-scrubbing solution), which is introduced at the top of the packing. This results in the highest possible efficiency. Since the solute concentration in the gas stream decreases as it rises through the column, there is fresh solvent constantly available for contact. This provides the maximum average driving force for the mass transfer process throughout the packed bed.

Mist eliminators also play an important role in absorbers. Mist eliminators are used to remove liquid droplets entrained in the gas stream. Ease of separation depends

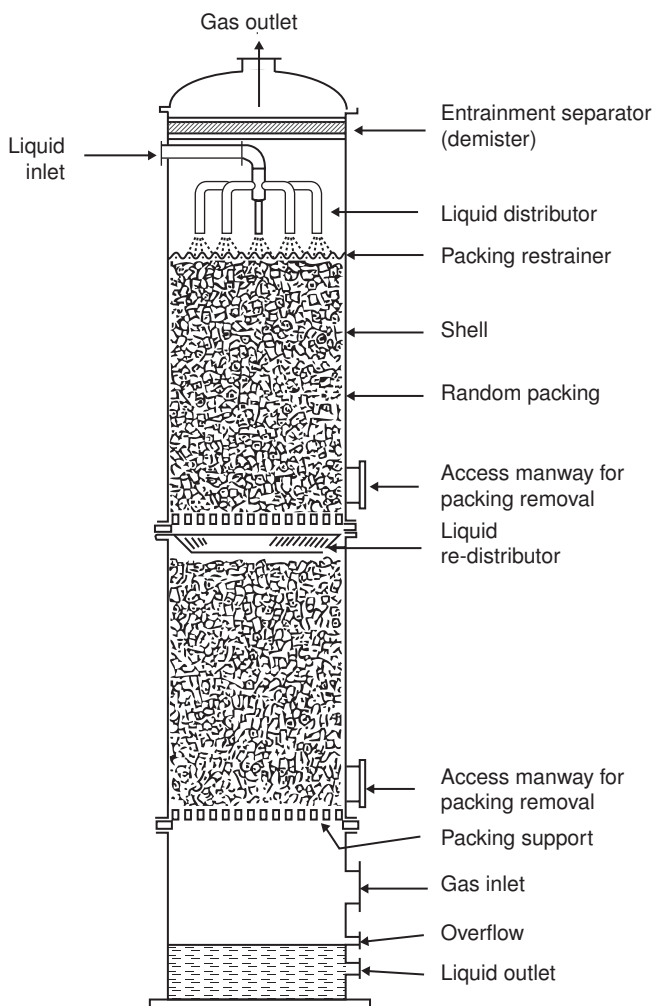


Figure 10.1 Typical countercurrent packed column.

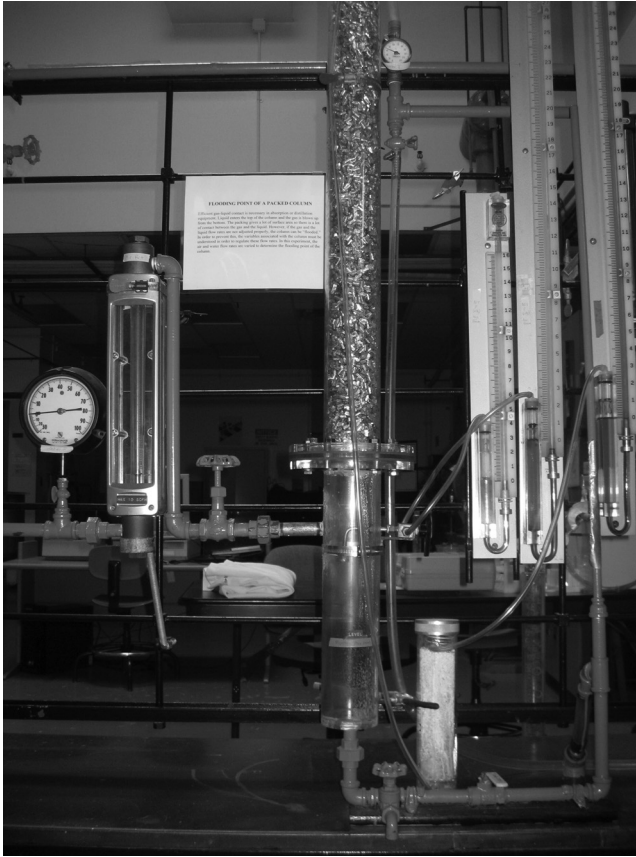


Figure 10.2 Absorption column.

on the size of the droplets. Droplets formed from liquids are usually large—up to hundreds of microns in diameter. However, drops formed in condensation or chemical reactions may be less than one micron in size. Entrainment removal (mist separation) is possible by a number of methods including the following:

- 1 Knitted wire or plastic mesh
- 2 Swirl vanes or zigzag vanes
- 3 Cyclones
- 4 Gravity settling chambers
- 5 Units in which the gas is forced to make a 180° turn
- 6 Additional packing above the packed bed

One of the simplest and most efficient means of mist separation is to use a porous blanket of knitted wire or plastic mesh. For most processes, the pressure drop across these mist eliminators range from 0.1 to 1.0 inches of water, depending on vapor and liquid

flowrates and the size of the eliminators. The efficiency of separation is generally high—usually 90% or better.

The packing is the heart of this type of equipment. Its proper selection entails an understanding of packing operational characteristics and the effect of performance of the points of significant physical difference between the various types. The main points to be considered in choosing the column packing include:

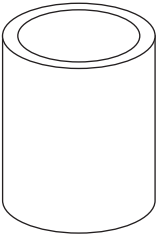

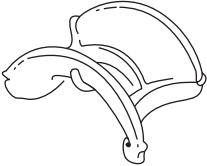
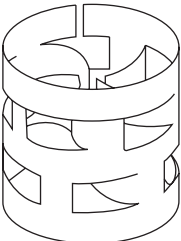
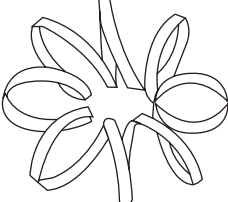
- 1 Durability and corrosion resistance (the packing should be chemically inert to the fluids being processed)
- 2 Free space per unit volume of packed space (this controls the liquor holdup in the column as well as the pressure drop across it; ordinarily, the fractional void volume, or fraction of free space, in the packed bed should be large)
- 3 Wetted surface area per unit volume of packed space. (This is very important since it determines the interfacial surface between liquid and gas; it is rarely equal to the actual geometric surface since the packing is usually not completely wetted by the fluid.)
- 4 Resistance to the flow of gas (this effects the pressure drop over the column)
- 5 Packing stability and structural strength to permit easy handling and installation
- 6 Weight per unit volume of packed space
- 7 Cost per unit area of packed space

Table 10.1 illustrates some of the various types and applications of the different column packings available.

One additional distinction should also be made: the difference between random and stacked (structured) packings. Random packings are those that are simply dumped into the column during installation and allowed to fall at random. It is the most common method of packing installation. During installation prior to pouring the packing into the column, the column may first be filled with water. This prevents breakage of the more fragile packing by reducing the velocity of the fall. The fall should be as gentle as possible since broken packing tightens the bed and increases the pressure drop. Stacked packing, on the other hand, is specially laid out and stacked by hand, making it a tedious operation and rather costly; it is avoided where possible except for the initial layers on supports. Liquid distributed in this latter system usually flows straight down through the packing immediately adjacent to the point of contact.

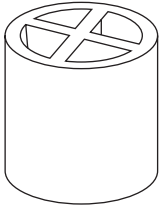
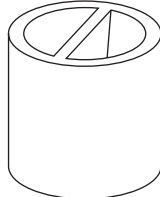
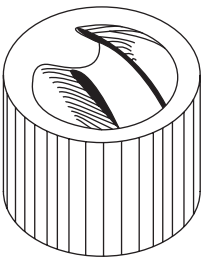
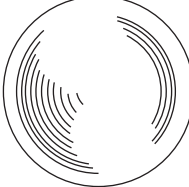
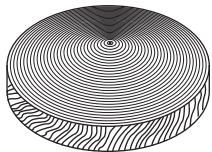
The aforementioned liquid distribution plays an important role in the efficient operation of the packed column. A good packing from a process viewpoint can be reduced in effectiveness by poor liquid distribution across the top of its upper surface. Poor distribution reduces the effective wetted packing area and promotes liquid channeling. The final selection of the mechanism of distributing the liquid across the packing depends on the size of the column, type of packing, tendency of the packing to divert liquid to column walls, and materials of construction for distribution. For stacked packing, the liquid usually has little tendency to cross distribute and thus moves down the column in the cross sectional area that it enters. In the dumped condition, most liquids follow a conical distribution down the column with the apex of the cone at the liquid impingement point. For uniform liquid flow and reduced channeling

Table 10.1 Some Typical Packings and Applications

Packing	Application features
Raschig rings 	Originally, the most popular type, usually cheaper per unit cost but sometimes less efficient than others; available in widest variety of materials to fit service: very sound structurally; usually packed by dumping wet or dry, with larger 4- to 6-in sizes sometimes hand-stacked; wall thickness varies between manufacturers; available surface changes with wall thickness; produce considerable side thrust on tower; usually has more internal liquid channeling and directs more liquid to walls of column.
Berl saddles 	More efficient than Raschig rings in most applications, but more costly; packing nests together and creates "tight" spots in bed that promotes channeling but not as much as Raschig rings; do not produce much side thrust and have lower unit pressure drops with higher flooding points than Raschig rings; easier to break in bed than Raschig rings.
Intalox saddles 	One of the most efficient packings, but more costly; very little tendency or ability to nest and block areas of bed; higher flooding limits and lower pressure drop than Raschig rings or Berl saddles; easier to break in bed than Raschig rings.
Pall rings 	Lower pressure drop (less than half) than Raschig rings; higher flooding limit; good liquid distribution; high capacity; considerable side thrust on column wall; available in metal, plastic, and ceramic.
Spiral rings 	Usually installed as stacked, taking advantage of internal whirl of gas-liquid and offering extra contact surface over Raschig rings.

(Continued)

TABLE 10.1 *Continued*

Packing	Application features
Teller rosette (Tellerette)	Available in plastic; lower pressure drops; higher flooding limits than Raschig rings or Berl saddles; very low unit weight; low side thrust; relatively expensive.
	
Cross-partition rings	Usually stacked as first layers on support grids for smaller packing above; pressure drop relatively low; channeling reduced for comparative stacked packings; no side wall thrust.
	
Lessing rings	Not much performance data available, but in general slightly better than Raschig rings; pressure drop slightly higher; high side-wall thrust.
	
Ceramic balls	Tend to fluidize in certain operating ranges, self-cleaning, uniform bed structure, higher pressure drop, and better contact efficiency than Raschig rings; high side thrust; not much commercial data.
	
Goodloe packing and wire mesh packing	Available in metal only, used in large and small columns for distillation, absorption, scrubbing, liquid extraction; high efficiency; low pressure drop.
	

of gas and liquid, the introduction of the liquid onto the packed bed must be as uniform as possible. Any impingement of the liquid on the wall of the column should be redistributed after a bed depth of approximately three column diameters for Raschig rings and five to ten column diameters for saddle packings. As a guide, Raschig rings usually have a maximum 10–15 ft of packing per section, while saddle packing can use a maximum of 12–20 ft. As a general rule of thumb, however, the liquid should be redistributed every 10 ft of packed height. The redistribution brings the liquid off the wall and directs it toward the center area of the column for a new start of distribution and contact in the next lower section.

Occasionally, cocurrent flow may be used where the gas stream and solvent both enter the top of the column. Initially, there is a very high rate of absorption that constantly decreases until, with an infinitely tall column, the gas and liquid leave in equilibrium and effectively operate as one theoretical stage. In this case, high gas and liquid rates are possible since the pressure drop tends to be rather low. However, these columns are efficient only when large driving forces are available (e.g., with very soluble gases or acid scrubbing in caustic media). The design for this case utilizes minimum column diameter because of the low pressure drop and nonflooding characteristics.

Packed columns may also operate in a crossflow mode (see Fig. 10.3) where the air stream moves horizontally through the packed bed and is irrigated by the scrubbing liquid which flows vertically down through the packing. Crossflow designs are characterized by low water consumption and fairly high air flow capacity at a low pressure drop. Where highly soluble gases are to be recovered, the crossflow packed scrubber has several advantages over the countercurrent scrubber. For example, when operating with the same liquid and gas mass flow rates, a crossflow scrubber has a lower pressure drop. Besides reducing water consumption drastically, the crossflow principle also reduces pump and fan motor sizes. Other advantages include less piping, less plugging from solids dropout at the packing support plate, and the possible use of higher gas and

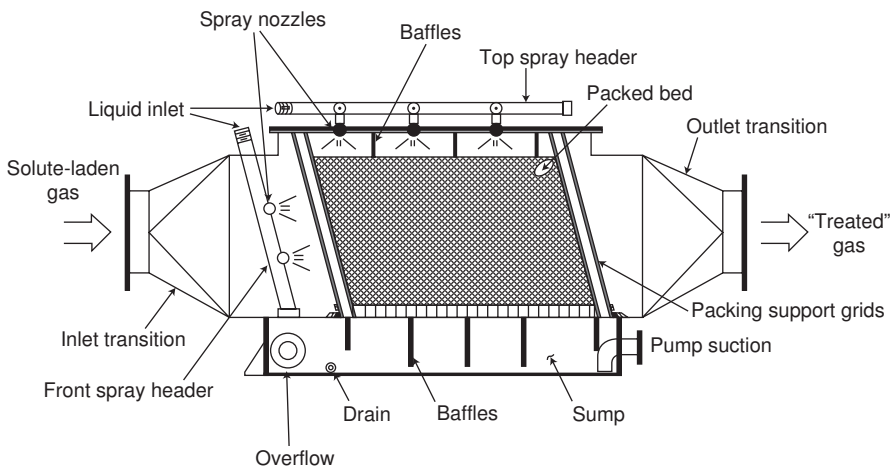


Figure 10.3 Cross flow operation in a packed column.

liquid rates because of the extremely low pressure drop. On the other hand, liquid entrainment from these systems is rather high and mist eliminators are usually required downstream.

Packed columns are characterized by a number of features to which their widespread popularity may be attributed.

- 1 Minimum structure—the packed column usually needs only a packing support and liquid distributor approximately every 10 feet along its height
- 2 Versatility—the packing material can be changed by simply discarding it and replacing it with a type providing better efficiency
- 3 Corrosive-fluids handling—ceramic packing is used and may be preferable to metal or plastic because of its corrosion resistance. When packing does deteriorate, it is quickly and easily replaced; it is also preferred when handling hot combustion gases
- 4 Low pressure drop—unless operated at very high liquid rates where the liquid becomes the continuous phase as the flowing films thicken and merge, the pressure drop per lineal foot of packed height is relatively low
- 5 Range of operation—although efficiency varies with gas and liquid feed rates, the range of operation is relatively broad
- 6 Low investment—when plastic packings are satisfactory or when the columns are less than about 3 or 4 feet in diameter, cost is relatively low

Plate Columns

Plate columns (also commonly referred to as “tray columns”) are essentially vertical cylinders in which the liquid and gas are contacted in stepwise fashion (staged operation) on plates or “traps,” as shown schematically for one type in Figure 10.4. The liquid enters at the top and flows downward via gravity. On the way, it flows across each plate and through a downspout to the plate below. The gas passes upward through openings of one sort or another in the plate, then bubbles through the liquid to form a froth, disengages from the froth, and passes on to the next plate above. The overall effect is a multiple countercurrent contact of gas and liquid. Each plate of the column is a stage since the fluids on the plate are brought into intimate contact, interface diffusion occurs, and the fluids are separated. The number of theoretical plates (or stages) is dependent on the difficulty of the separation to be carried out and is determined solely from material balances and equilibrium considerations. The diameter of the column, on the other hand, depends on the quantities of liquid and gas flowing through the column per unit time. The actual number of plates required for a given separation is greater than the theoretical number because of plate inefficiencies. To achieve high plate efficiencies, the contact time between the gas and liquid on each plate should be high so as to permit mass transfer to occur, the interfacial surface between phases must be as large as possible, and a relatively high degree of turbulence is required to obtain high mass transfer coefficients. In order to increase contact time, the liquid pool height on each plate should be deep so that the gas bubbles will require

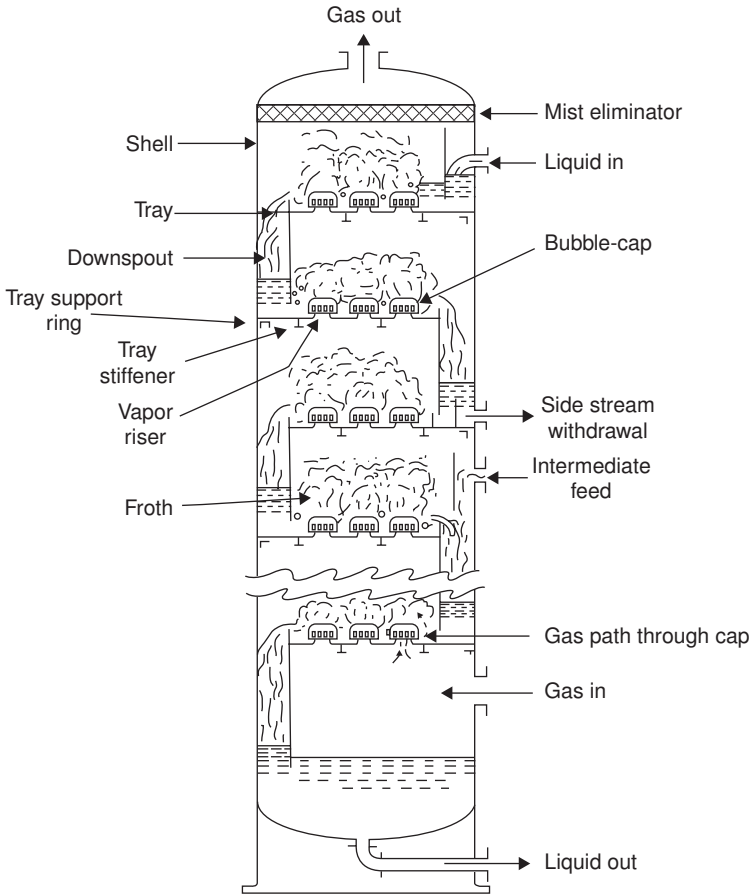


Figure 10.4 Typical bubble-cap plate column.

a relatively long time to rise through the liquid. On the other hand, great depths of liquid on the plates, although leading to high plate efficiencies, results in a higher pressure drop per plate. Relatively high gas velocities are also preferred for high plate efficiencies. This results in the gas being thoroughly dispersed into the liquid and causes froth formation, which provides large interfacial surface areas. High gas velocities, although providing good vapor–liquid contact, may lead to excessive entrainment accompanied by high pressure drop. Hence, the various arrangements and dimensions chosen for the design of plate columns are usually those which experience has proven to provide reasonably good compromises. Additional information on the design of plate columns can be found in Chapter 9 and in a later section.

The particular plate selection and its design can materially affect the performance of a given absorption operation. Each plate should be designed so as to provide as efficient a contact between the vapor and liquid as possible, within reasonable economic limits. The principal types of plates encountered are discussed .

In bubble-cap plates (as discussed in the previous chapter), the vapor rises up through a “riser” into the bubble-cap, out through the slots as bubbles, and into the surrounding liquid on the plates. Figure 10.5 demonstrates the liquid–vapor action for a bubble-cap plate. The bubble-cap plate design was once the most favored of plate designs.

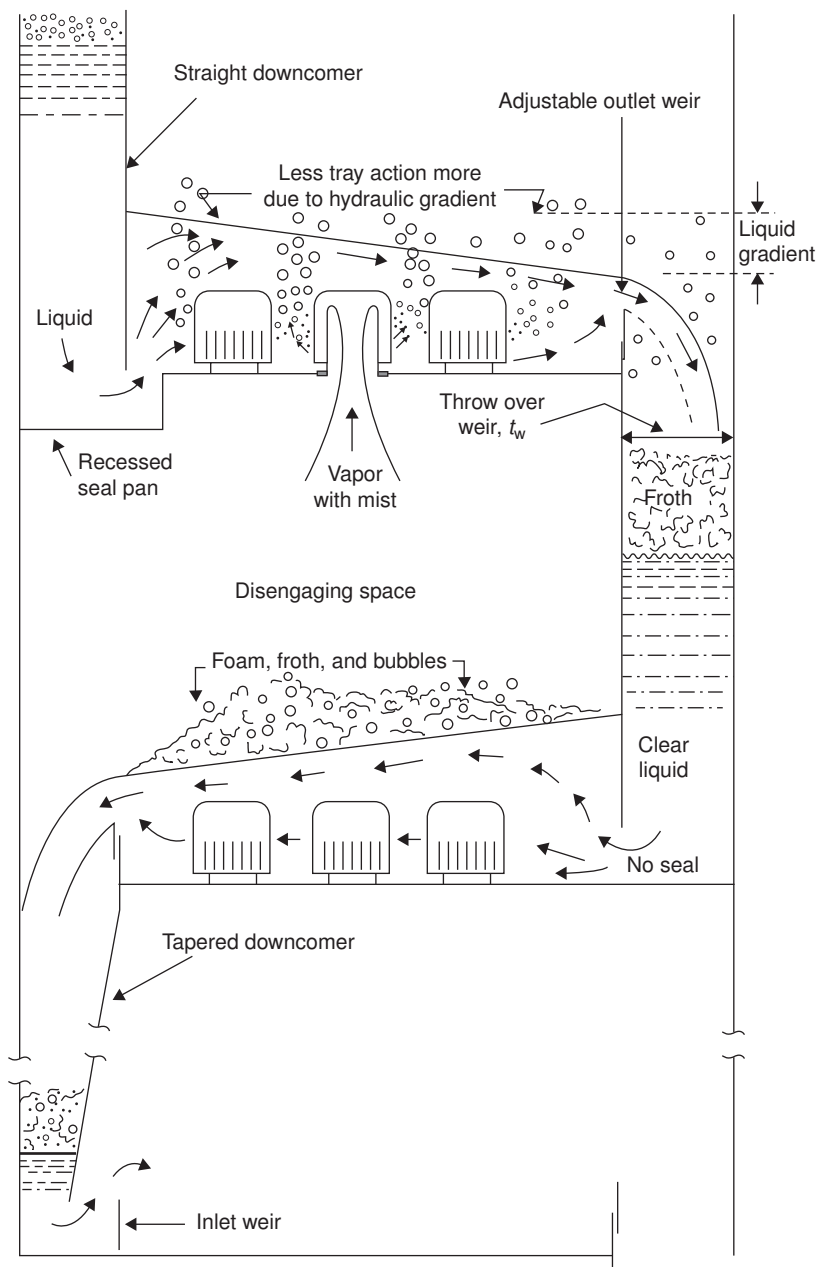


Figure 10.5 Bubble-cap plate schematic-dynamic operation.

In sieve or perforated plates, the vapor rises through small holes (usually $\frac{1}{8}$ to 1 inch in diameter) in the plate floor and bubbles through the liquid in a fairly uniform manner. The perforated plate is made with or (occasionally) without the downcomer. With the downcomer, the liquid flows across the plate floor and over a weir (if used), then through the downcomer to the plate below. Figure 10.6 shows the operation

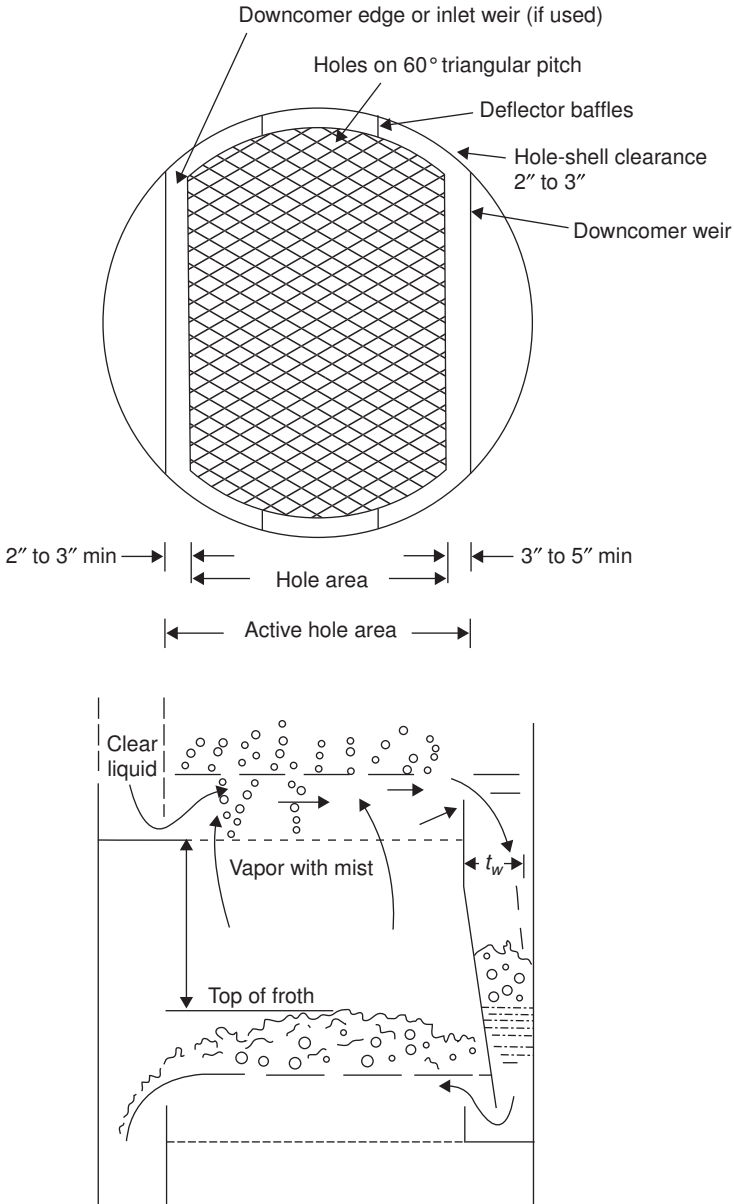


Figure 10.6 Sieve or perforated plate with downcomers.

schematically. These plates are generally not suitable for columns operating under variable load. Plate spacing in this case usually averages about 15 inches. At the same time that the vapor rises through the holes, the liquid head forces liquid counter-current through these holes and onto the plate below. Perforated plates have become the preferred choice in recent years.

DESIGN AND PERFORMANCE EQUATIONS—PACKED COLUMNS

Design and performance equations are provided for both packed and plate columns, and some overlap does exist. However, the emphasis is on packed columns since they are often the choice in absorption applications. On the other hand, plate columns, the preferred choice in distillation (see previous chapter), will receive attention in the next section.

For most absorption applications, sufficient information is either provided or available to enable one to completely describe the system through simple yet standard calculational procedures. These calculations generally involve the determination of three unknown system variables: the liquid rate, the column height (and corresponding pressure drop), and the column diameter. Each of the topics is treated sequentially below.

Liquid Rate

As described earlier, the equilibrium of interest in gas absorption is that between a relatively nonvolatile absorbing liquid (solvent) and a soluble gas (solute). For cases that follow Henry's law (see Chapter 6), Henry's law constant, m , is defined by

$$y^* = mx \quad (10.1)$$

The usual operating data to be determined or estimated for isothermal systems are the liquid rates and the terminal concentrations or mole fractions. An operating line, which describes operating conditions in a countercurrent flow column, is obtained by a mass (or a mole basis) balance around the column (as shown in Fig. 10.7). Note that the notation normally employed for the gas rate in absorption calculations is G , not V , as employed for the vapor rate in Chapter 9. The subscript m is often carried if the rate is based on moles. The overall mole balance is:

$$G_{m1} + L_{m2} = G_{m2} + L_{m1} \quad (10.2)$$

For component A , the mass (or mole) balance becomes

$$G_{m1}y_{A1} + L_{m2}x_{A2} = G_{m2}y_{A2} + L_{m1}x_{A1} \quad (10.3)$$

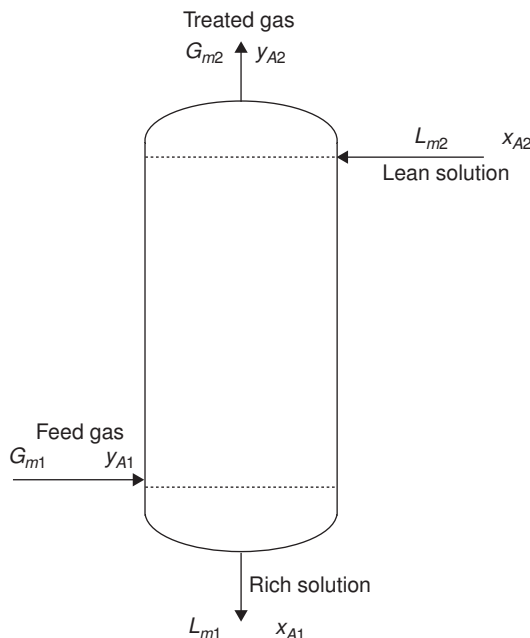


Figure 10.7 Mole balance; countercurrent flow.

Assuming $G_{m1} = G_{m2} = G_m$ and $L_{m1} = L_{m2} = L_m$ (reasonable for many applications where solute concentrations are reasonably small), then

$$G_m y_{A1} + L_m x_{A2} = G_m y_{A2} + L_m x_{A1}$$

or, on rearrangement,

$$\frac{L_m}{G_m} = \frac{y_{A1} - y_{A2}}{x_{A1} - x_{A2}} \quad (10.4)$$

This is the equation of a straight line known as the operating line. It has a slope of L_m/G_m on x, y coordinates and passes through the points (x_{A1}, y_{A1}) and (x_{A2}, y_{A2}) as indicated in Figure 10.8.

In the design of most absorption columns, the quantity of gas to be treated, G_m , the concentrations, y_{A1} and y_{A2} , and the composition of the entering liquid, x_{A2} , are ordinarily fixed by process requirements. However, the quantity of liquid solvent to be used is subject to some choice. This can be resolved by setting or obtaining a minimum liquid-to-gas ratio.

With reference to Figure 10.8, the operating line must pass through point A (top of column) and must terminate at the ordinate y_{A1} . If such a quantity of liquid is used to

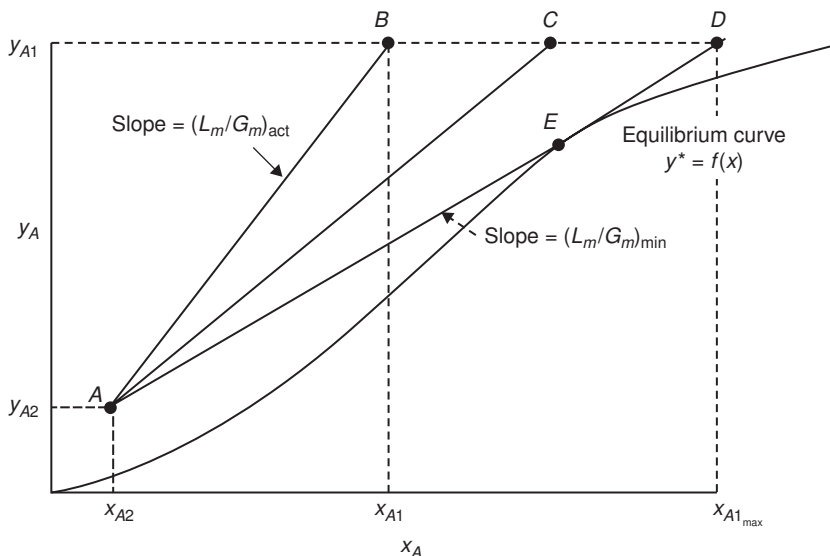


Figure 10.8 Operating and equilibrium lines.

produce operating line AB , the exiting liquid will have the composition x_{A1} . If less liquid is used, the exit liquid composition will clearly be greater, as at point C , but since the driving forces for mass transfer are less, the absorption is more difficult. The time of contact between the gas and liquid must then be greater and the absorber must be correspondingly taller.

The minimum liquid that can be used corresponds to the operating line AD , which has the greatest slope for any line touching the equilibrium curve (tangent to the curve at E). At point E , the mass transfer driving force is zero, the required contact time for the concentration change desired is infinite, and an infinitely tall column results. This then represents the minimum liquid-to-gas ratio. The importance of the minimum liquid-to-gas ratio lies in the fact that column operation is frequently specified as some factor of the minimum liquid-to-gas ratio. For example, a typical situation frequently encountered is that the actual operating line, $(L_m/G_m)_{act}$ is $1.5(L_m/G_m)_{min}$.

ILLUSTRATIVE EXAMPLE 10.1

Experimental data are provided for an absorption system to be used for scrubbing ammonia (NH_3) from air with water. The water rate is 300 lb/min and the gas rate is 250 lb/min at 72°F. The applicable equilibrium data for the ammonia-air system is shown in Table 10.2. The air to be scrubbed has 1.5% (mass basis) NH_3 at 72°F and 1 atm pressure and is to be vented with 95% of the ammonia recovered. The inlet scrubber water is ammonia-free.

- 1 Plot the equilibrium data in mole fraction units.
- 2 Perform the material balance and plot the operating line on the equilibrium plot.

Table 10.2 Ammonia Equilibrium Data I

Equilibrium partial pressure (mm Hg)	NH ₃ concentration (lb NH ₃ /100 lb H ₂ O)
3.4	0.5
7.4	1.0
9.1	1.2
12.0	1.6
15.3	2.0
19.4	2.5
23.5	3.0

SOLUTION

- 1 Employing the data provided in the problem statement, convert the equilibrium partial pressure data and the liquid concentration data to mole fractions as shown in Table 10.3. Plotting the mole fraction values on the graph in Figure 10.9 results in a straight line. The slope of the equilibrium line is approximately 1.0.
- 2 Convert the liquid and gas rates to lbmol/min, assuming that the molecular weights are approximately that of pure water and air:

$$L_m = L/18 = 300/18 = 16.67 \text{ lbmol/min}$$

$$G_m = G/29 = 250/29 = 8.62 \text{ lbmol/min}$$

Determine the inlet and outlet mole fractions for the gas, y_1 and y_2 , respectively (dropping the subscripts):

$$y_1 = \frac{1.5/17}{1.5/17 + 98.5/29} = 0.0253$$

Table 10.3 Ammonia Equilibrium Data II

Gas mole fraction, y	Liquid mole fraction, x
0.00447	0.0053
0.00973	0.0106
0.0120	0.0127
0.0158	0.0169
0.0201	0.0212
0.0255	0.0265
0.0309	0.0318

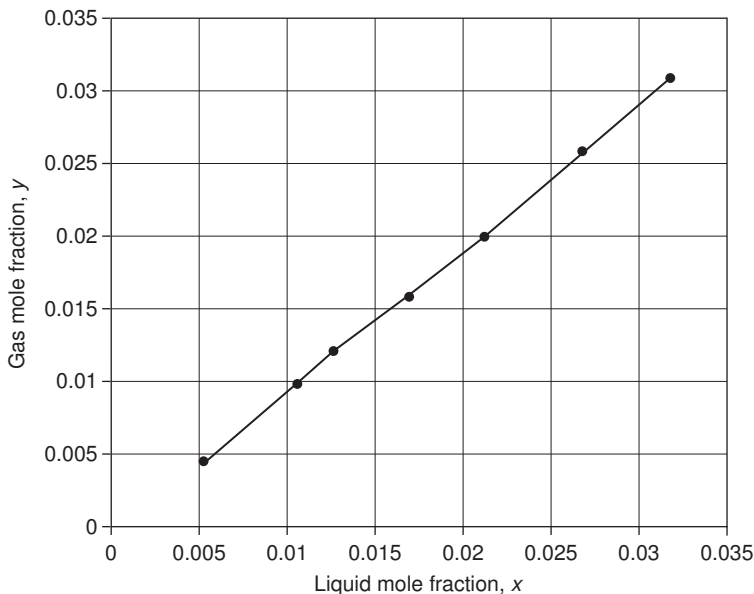


Figure 10.9 Ammonia equilibrium curve in water (72°F, 1 atm).

$$y_2 = \frac{(0.05)(1.5/17)}{(0.05)(1.5/17) + (98.5/29)} = 0.0013$$

The inlet liquid mole fraction x_2 is given as zero, and the describing equation for x_1 , the outlet liquid mole fraction, is

$$x_1 = (G_m/L_m)(y_1 - y_2) + x_2 \quad (10.4)$$

Substituting gives

$$x_1 = (8.62/16.67)(0.0253 - 0.0013) + 0 = 0.0124$$

One may now use the inlet and outlet mole fractions to plot the operating line on the graph in Figure 10.10.

The slope of the operating line is

$$\text{Slope} = (L_m/G_m)_{\text{act}} = \frac{0.0253 - 0.0013}{0.0124 - 0.0} = 1.935$$

The reader is left the exercise of calculation $(L_m/G_m)_{\text{min}}$. (The answer is slightly greater than 1.0). ■

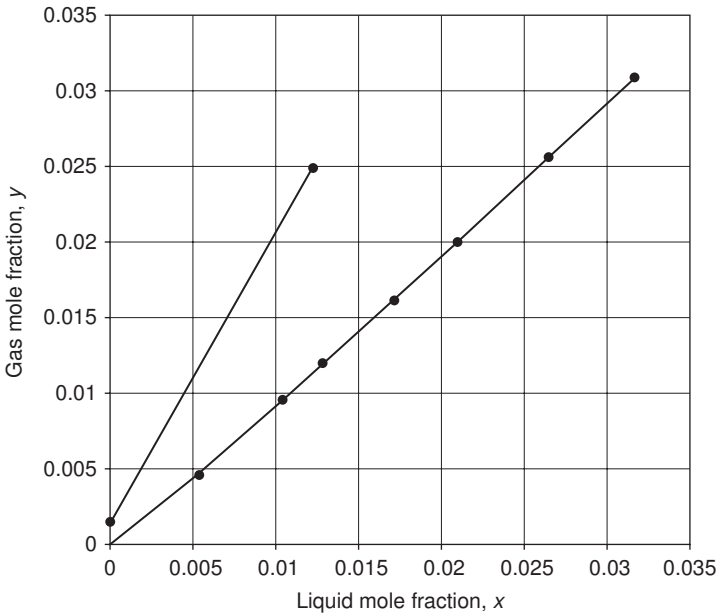


Figure 10.10 Operating line for Illustrative Example 10.1.

ILLUSTRATIVE EXAMPLE 10.2

Given the following information for a packed countercurrent gas scrubber, determine the liquid flux in $\text{lbmol}/\text{h} \cdot \text{ft}^2$.

Gas flux = $18 \text{ lbmol}/\text{h} \cdot \text{ft}^2$

The mole fractions of the solute in the inlet and outlet gas are 0.08 and 0.002, respectively

The mole fractions of the solute in the inlet and outlet liquid are 0.001 and 0.05, respectively.

SOLUTION: Applying a componential mole balance (dropping the subscripts once again) on the solute gives an equivalent form of Equation (10.4):

$$L_m(x_{\text{out}} - x_{\text{in}}) = G_m(y_{\text{in}} - y_{\text{out}}) \quad (10.4)$$

Substituting, one obtains

$$\begin{aligned} L_m(0.05 - 0.001) &= 18(0.08 - 0.002) \\ L_m &= 18(0.078)/0.049 \\ &= 28.7 \text{ lbmol}/\text{h} \cdot \text{ft}^2 \end{aligned}$$

■

ILLUSTRATIVE EXAMPLE 10.3

The EPA has conducted investigations on the exhaust streams at the Buonicore Chemical Company. To Buonicore's dismay, their hydrocarbon emissions are too high and must be reduced in order to continue operation. At present conditions, the exhaust stream contains 1.0% benzene at a total flow of 40,000 ft³/h. It has been determined that if the exhaust stream is reduced to 0.01% benzene, the company can continue operation.

An absorber that is currently out of commission will be used to absorb the benzene, employing a light wash oil as the absorbent. The pump on the liquid feed has a maximum liquid pumping rate of 50 ft³/h. Does this pump have enough capacity to do the job if the outlet benzene concentration in the wash oil cannot exceed 5.0%? Operating data are provided below:

Gas temperature = 100°F

Liquid temperature = 60°F

Light oil molecular weight (MW) = 156 lb/lbmol

Light oil density = 55.1 lb/ft³

Actual liquid flow rate = (1.5) $L_{m,\min}$

SOLUTION: Find the gas molar flow rate G_m by applying the ideal gas law:

$$\begin{aligned} G_m &= (40,000 \text{ ft}^3/\text{hr})(\text{lbmol}/379 \text{ ft}^3)[(460 + 100)/(460 + 60)] \\ &= 114 \text{ lbmol/hr} \end{aligned}$$

Calculate the minimum liquid rate $L_{m,\min}$ by applying Equation (10.4):

$$\begin{aligned} \frac{L_m}{G_m} &= \frac{y_{\text{in}} - y_{\text{out}}}{x_{\text{out}} - x_{\text{in}}} \\ L_{m,\min} &= 114 \left(\frac{0.01 - 0.0001}{0.05 - 0} \right) \\ &= 22.6 \text{ lbmol/h} \end{aligned}$$

The actual operating liquid molar flow rate is

$$\begin{aligned} L_m &= 1.5(22.6) \\ &= 33.9 \text{ lbmol/h} \end{aligned}$$

The required liquid mass flow rate is

$$\begin{aligned} L &= 33.9(156 \text{ lb/lbmol}) \\ &= 5288 \text{ lb/h} \end{aligned}$$

The liquid volumetric flow rate is then

$$\begin{aligned} q_L &= 5288/55.1 \\ &= 96.0 \text{ ft}^3/\text{h} \end{aligned}$$

The present pump does *not* have enough capacity. ■

Column Diameter

Consider a packed column operating at a given liquid rate and the gas rate is then gradually increased. After a certain point, the gas rate is so high that the drag on the liquid is sufficient to keep the liquid from flowing freely down the column. Liquid begins to accumulate and tends to block the entire cross section for flow (a process referred to as *loading*). This, of course, increases both the pressure drop and prevents the packing from mixing the gas and liquid effectively, and ultimately some liquid is even carried back up the column. This undesirable condition, known as *flooding*, occurs fairly abruptly, and the superficial gas velocity at which it occurs is called the *flooding velocity*. The calculation of column diameter is usually based on flooding considerations, with the usual operating range being taken as 50–75% of the flooding rate.

One of the more commonly used correlations is U.S. Stoneware’s⁽³⁾ generalized pressure drop correlation, as presented in Figure 10.11. The procedure to determine the column diameter is as follows:

- 1 Calculate the abscissa, $(L/G)(\rho_G/\rho_L)^{0.5}$; mass basis for all terms
- 2 Proceed to the flooding line and read the ordinate (design parameter)
- 3 Solve the ordinate equation for G_f at flooding
- 4 Calculate the column cross-sectional area, S , for the fraction of flooding velocity chosen for operation, f , by the equation:

$$S = \frac{W}{fG_f} \tag{10.5}$$

where W (\dot{m}) is the mass flow rate of the gas in lb/s and S is the area in ft².

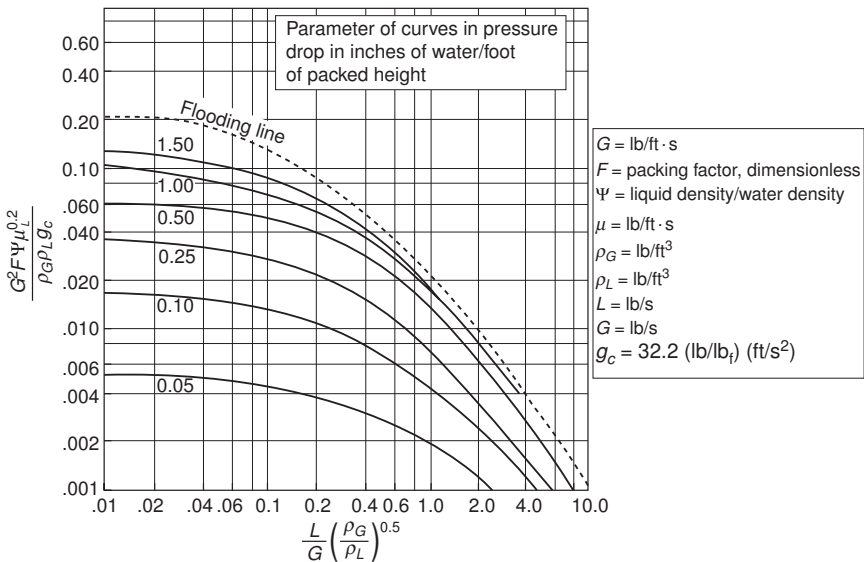


Figure 10.11 Generalized pressure drop correlation to estimate column diameter.

5 The diameter of the column is then determined by

$$D = \left(\frac{4}{\pi} S \right)^{0.5} = 1.13S^{0.5}; \text{ ft} \quad (10.6)$$

Note that the proper units, as designated in the correlation, must be used as the plot is not dimensionless. The flooding rate is usually evaluated using total flows of the phases at the bottom of the column where they are at their highest value. The pressure drop may be evaluated directly from Figure 10.11 using a revised ordinate that contains the actual, not flooding, value of G .

Chen⁽⁴⁾ developed the following equation from which the tower diameter can easily be obtained:

$$D = 16.28 \left(\frac{W}{\phi L} \right)^{0.5} \left(\frac{\rho_L}{\rho_G} \right)^{0.25} \quad (10.7)$$

where

$$\log_{10} \phi = 32.5496 - 4.1288 \log_{10} \left(\frac{L^2 A_v \mu_L^{0.2}}{\rho L^2 \varepsilon^3} \right)^{0.5} \quad (10.8)$$

where (employing Chen's notation) A_v is the specific surface area of dry packing (ft^2/ft^3 packed column), L is the liquid flux ($\text{gal}/\text{min} \cdot \text{ft}^2$ of superficial tower cross section), W is the mass rate of flow of gas (lb/h), ε is the void fraction, μ_L is the liquid viscosity (cP), and the density terms are in lb/ft^3 .

ILLUSTRATIVE EXAMPLE 10.4

A packed column is used to absorb a toxic pollutant from a gas stream. From the data given below, calculate the height of packing and column diameter. The unit operates at 50% of the flooding gas mass velocity, the actual liquid flow rate is 40% more than the minimum, and 95% of the pollutant is to be collected. Employ the generalized correlation provided in Figure 10.11 to estimate the column diameter.

Gas mass flow rate = 3500 lb/h

Pollutant concentration in inlet gas stream = 1.1 mol%

Scrubbing liquid = pure water

Packing type = 1-inch Raschig rings; packing factor $F = 160$

H_{OG} of the column = 2.5 ft

Henry's law constant $m = 0.98$

Density of gas (air) = 0.075 lb/ft^3

Density of water = 62.4 lb/ft^3

Viscosity of water = 1.8 cP

SOLUTION: First calculate the equilibrium outlet concentration x_1^* at $y_1 = 0.011$:

$$x_1^* = y_1/m \quad (10.1)$$

Substituting

$$\begin{aligned} x_1^* &= 0.011/0.98 \\ &= 0.0112 \end{aligned}$$

Determine y_2 for 95% removal:

$$\begin{aligned} y_2 &= \frac{(0.05)y_1}{(1 - y_1) + (0.05)y_1} \\ &= \frac{(0.05)(0.011)}{(1 - 0.011) + (0.05)(0.011)} \\ &= 5.56 \times 10^{-4} \end{aligned}$$

The minimum ratio of molar liquid flow rate to molar gas flow rate, $(L_m/G_m)_{\min}$, is determined by a material balance employing the equilibrium value at the top of the column, i.e., x^* rather than x :

$$\begin{aligned} \left(\frac{L_m}{G_m}\right)_{\min} &= \frac{y_1 - y_2}{x_1^* - x_2} \quad (10.4) \\ &= \frac{0.011 - 5.56 \times 10^{-4}}{0.0112 - 0} \\ &= 0.933 \end{aligned}$$

The actual ratio of molar liquid flow rate to molar gas flow rate L_m/G_m is

$$\begin{aligned} (L_m/G_m)_{act} &= L_m/G_m = (1.40)(L_m/G_m)_{\min} \\ &= (1.40)(0.933) \\ &= 1.306 \end{aligned}$$

In addition

$$(mG_m)/L_m = (0.98)/(1.306) = 0.7504$$

To determine the diameter of the packed column, the ordinate of Figure 10.11 is first calculated:

$$\begin{aligned} \left(\frac{L}{G}\right)\left(\frac{\rho_G}{\rho_L}\right)^{0.5} &= \left(\frac{L_m}{G_m}\right)\left(\frac{18}{29}\right)\left(\frac{\rho_G}{\rho_L}\right)^{0.5} = (1.306)\left(\frac{18}{29}\right)\left(\frac{0.075}{62.4}\right)^{0.5} \\ &= 0.0281 \end{aligned}$$

The value of the abscissa at the flooding line is determined from the same figure:

$$\frac{G^2 F \Psi \mu_L^{0.2}}{\rho_L \rho_G g_c} = 0.21;$$

The flooding gas mass flux G_f in $\text{lb}/\text{ft}^2 \cdot \text{s}$ is

$$\begin{aligned} G_f &= \left(\frac{0.21 \rho_L \rho_G g_c}{F \Psi \mu_L^{0.2}}\right)^{1/2} = \left(\frac{(0.21)(62.4)(0.075)(32.2)}{(160)(1.0)(1.8)^{0.2}}\right)^{1/2} \\ &= 0.419 \text{ lb}/\text{ft}^2 \cdot \text{s} \end{aligned}$$

Apply Equation (10.5) to determine the column cross-sectional area.

$$S = \frac{W}{FG_F} = \frac{(3500/3600)}{(0.5)(0.419)} = 4.64 \text{ ft}^2$$

Apply Equation (10.6) to calculate the column diameter.

$$D = \left(\frac{4}{\pi} S\right)^{0.5} = 1.13 S^{0.5} = (1.13)(4.64)^{0.5} = 2.43 \text{ ft}$$

Column Height

The column height may be estimated from

$$Z = N_{OG} H_{OG} \quad (10.9)$$

where N_{OG} is the number of overall transfer units, H_{OG} is the height of a single transfer unit and Z is the height of the column packing. In most design applications, the number of transfer units (N_{OG}) is obtained experimentally or calculated using any of the methods to be explained later in this section. The height of a transfer unit (H_{OG}) is also usually determined experimentally for the system under consideration. Information on many different systems using various types of packings has been compiled by the manufacturers of gas absorption equipment and should be consulted prior to design. The data may be in the form of graphs depicting, for a specific system and packing, the H_{OG} vs the gas mass flux ($\text{lb/h} \cdot \text{ft}^2$) with the liquid rate as a parameter. The packing height Z is then simply the product of the H_{OG} and the N_{OG} . Although there are many different approaches to determine the column height, the H_{OG} - N_{OG} approach is the simplest and presently the most used, with the H_{OG} usually being obtained from the manufacturer. Details on N_{OG} follow.

In many operations, the constituent to be absorbed (e.g., HCl) is in the very dilute range. For this condition

$$N_{OG} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} \quad (10.10)$$

If the operating line and equilibrium line are both parallel and straight:

$$N_{OG} = \frac{y_1 - y_2}{y - y^*} \quad (10.11)$$

If the operating line and equilibrium line are just straight (and not necessarily parallel)

$$N_{OG} = \frac{y_1 - y_2}{(y - y^*)_{lm}}; \quad (y - y^*)_{lm} = \frac{(y - y^*)_1 - (y - y^*)_2}{\ln[(y - y^*)_1 / (y - y^*)_2]} \quad (10.12)$$

However, one can show that if Henry’s law applies, the number of transfer units is given by Coburn’s equation

$$N_{OG} = \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{1 - \frac{1}{A}} \tag{10.13}$$

where

$$A = \frac{L_m}{mG_m} \tag{10.14}$$

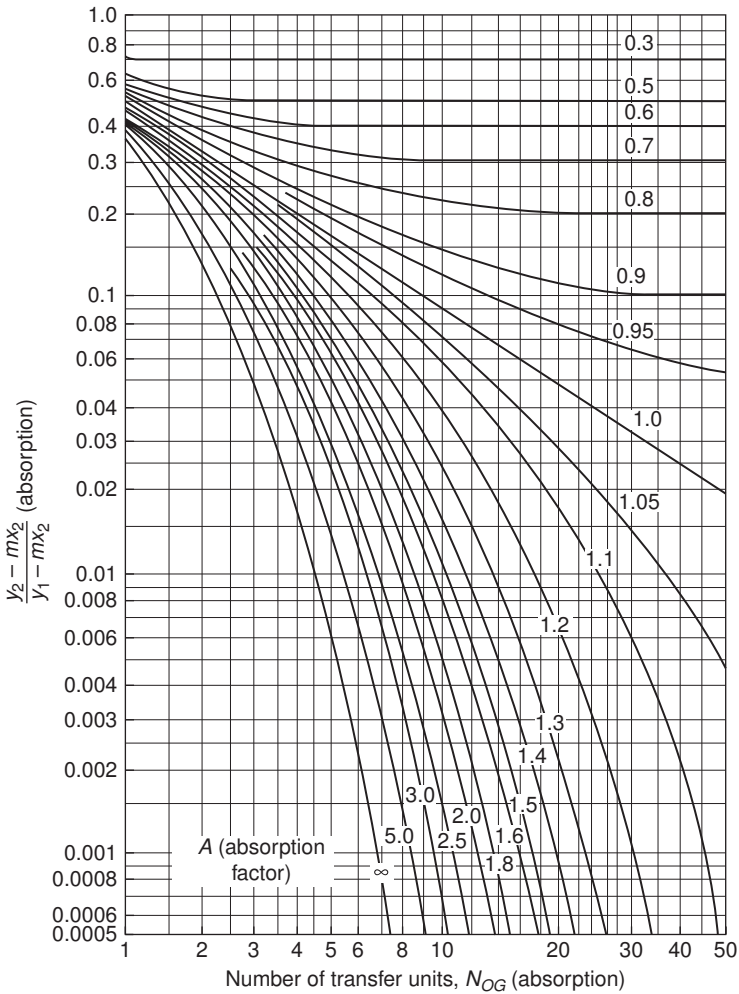


Figure 10.12 N_{OG} for absorption column with constant absorption factor.

and A is defined once again the absorption factor and m is the slope of the equilibrium curve. The solution to this equation can be conveniently found graphically from Figure 10.12. However, note that the flow rates L_m and G_m are based on moles in Equation (10.14). If the gas is highly soluble in the liquid and/or reacts with the liquid, Theodore⁽⁵⁾ has shown that

$$N_{OG} = \ln\left(\frac{y_1}{y_2}\right) \quad (10.15)$$

If the operating line and/or equilibrium line are curved, the integral above in Equation (10.10) should be evaluated.

Qualitatively, the height of a transfer unit is a measure of the height of a contactor required to affect a standard separation, and it is a function of the gas flow rate, the liquid flow rate, the type of packing, and the chemistry of the system. As indicated above, experimental values for H_{OG} are generally available in the literature or from vendors.^(1,5)

ILLUSTRATIVE EXAMPLE 10.5

When a gas is highly soluble, the number of overall gas transfer units N_{OG} in a packed tower is given by

$$N_{OG} = \ln\left(\frac{y_1}{y_2}\right) \quad (10.15)$$

Calculate N_{OG} if $y_1 = 200$ ppm and $y_2 = 0.5$ ppm.

SOLUTION: Substituting in Equation (10.15) yields

$$\begin{aligned} N_{OG} &= \ln(200/0.5) \\ &= \ln(400) \\ &= 5.99 \end{aligned} \quad \blacksquare$$

ILLUSTRATIVE EXAMPLE 10.6

A steel pickling operation emits 300 ppm HCl with peak values of 500 ppm, 15% of the time. The air flow is a constant 25,000 acfm at 75°F and 1 atm. Only sketchy information was submitted with a scrubber permit application for a spray tower. You are requested, as a regulatory official, to determine if the spray unit is satisfactory.

System information:

Emission limit = 25 ppm HCl

Maximum gas velocity allowed through the tower = 3 ft/s

Number of sprays = 6

Diameter of the tower = 14 ft

The plans show a countercurrent water spray tower. For a very soluble gas (Henry's law constant is approximately zero), and the number of transfer units (N_{OG}) can be determined by Equation (10.15), i.e.,⁽⁵⁾

$$N_{OG} = \ln\left(\frac{y_1}{y_2}\right)$$

where y_1 is the concentration of inlet gas and y_2 is the concentration of outlet gas. In a spray tower, the number of transfer units (N_{OG}) for the first (or top) spray is about 0.7. Each lower spray will have only about 60% of the N_{OG} of the spray above it. The final spray, if placed in the inlet duct, has a N_{OG} of 0.5. The spray sections of a tower are normally spaced at three foot intervals. The inlet duct spray adds no height to the column.

SOLUTION: Calculate the gas superficial velocity through the tower in ft/s:

$$\begin{aligned} v &= \frac{q}{\pi D^2/4}; \quad q = \text{volumetric flow rate} \\ &= \frac{25,000}{\pi(14)^2/4} = 162.4 \text{ ft/min} = 2.7 \text{ ft/s} \end{aligned}$$

The gas velocity meets requirements since it is less than 3 ft/s.

Calculate the number of transfer units required to meet the regulation under worst case conditions:

$$\begin{aligned} N_{OG} &= \ln\left(\frac{y_1}{y_2}\right) \quad (10.15) \\ &= \ln\left(\frac{500 \times 10^{-6}}{25 \times 10^{-6}}\right) = 3.0 \end{aligned}$$

Determine the total number of transfer units provided by a tower with six spray sections. The result is given in Table 10.4. Note that this value is below the required value of 3.0. The spray unit is therefore *not* satisfactory.

Using the total N_{OG} from Table 10.4, calculate the outlet concentration of gas:

$$\begin{aligned} \frac{y_1}{y_2} &= \exp(N_{OG}) = \exp(2.114) = 8.28 \\ y_2 &= \frac{500}{8.28} = 60.4 \text{ ppm} \end{aligned}$$

Table 10.4 N_{OG} Calculation for Illustrative Example 10.6

Spray section		N_{OG}
Top	0.7	0.7
2nd	(0.7)(0.6) =	0.42
3rd	(0.42)(0.6) =	0.252
4th	(0.252)(0.6) =	0.1512
5th	(0.1512)(0.6) =	0.0907
Inlet	(Given)	0.5
Total		2.114



ILLUSTRATIVE EXAMPLE 10.7

Solve the following two problems:

- 1 Find the number of overall gas transfer units (N_{OG}) in a packed tower required to recover 90% of a gas in an inlet air stream containing 10 mole percent (mol%) solute using pure water at a rate 20% greater than the minimum rate. Assume $m = 1.485$.
- 2 How many N_{OG} values would be required if, instead of pure water, water containing 0.1, 0.3, 0.5, and 0.65 mol% (mole percent) of the solute in the gas were used instead?

SOLUTION

- 1 Calculations can be performed on a mole fraction or solute-free mole fraction basis. Since $y_1 = 0.1$, the solute-free mole fraction is

$$\begin{aligned} Y_1 &= y_1/(1 - y_1) \\ &= 0.1/(1 - 0.1) \\ &= 0.1111 \end{aligned}$$

In addition,

$$\begin{aligned} y_2 &= (0.1)(0.1111) \\ &= 0.0111 \end{aligned}$$

and

$$\begin{aligned} Y_2 &= 0.0111/(1 - 0.0111) \\ &= 0.0112 \end{aligned}$$

For the minimum rate, one obtains

$$\begin{aligned} X_1^* &= Y_1/1.485 \\ &= 0.1111/1.485 \\ &= 0.0748 \end{aligned}$$

The minimum liquid-to-gas ratio is then

$$\begin{aligned} \left(\frac{L_m}{G_m}\right)_{\min} &= \frac{0.1111 - 0.0112}{X_1^* - X_0} \\ &= \frac{0.1111 - 0.0112}{0.0748 - 0} \\ &= 1.34 \end{aligned}$$

The actual ratio is

$$\begin{aligned} \left(\frac{L_m}{G_m}\right)_{\text{act}} &= (1.2)(1.34) \\ &= 1.60 \end{aligned}$$

Since

$$A = \frac{L_m}{mG_m} \quad (10.14)$$

substituting gives

$$\begin{aligned} A &= 1.60/1.485 \\ &= 1.08; \quad 1/A = 0.928 \end{aligned}$$

and

$$\begin{aligned} \frac{y_2}{y_1} &\simeq \frac{Y_2}{Y_1} \\ &= 0.111; \quad \frac{y_1}{y_2} = 9.01 \end{aligned}$$

One may employ either Equation (10.13) or Figure 10.12. From Figure 10.12

$$N_{OG} \simeq 6.0$$

From Equation (10.13):

$$\begin{aligned} N_{OG} &= \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{1 - (1/A)} \\ &= \frac{\ln \left[\left(\frac{0.1}{0.0111} \right) (0.072) + 0.928 \right]}{0.072} \\ &= 6.32 \end{aligned}$$

2 For this condition

$$x_2 = X_2 = 0.001$$

Once again

$$\begin{aligned} \left(\frac{L_m}{G_m} \right)_{\min} &= \frac{0.1111 - 0.0111}{0.0748 - 0.001} \\ &= 1.35 \end{aligned}$$

and

$$\begin{aligned} \left(\frac{L_m}{G_m} \right)_{\text{act}} &= (1.2)(1.35) \\ &= 1.63 \end{aligned}$$

If Equation (10.13) is used, then

$$\begin{aligned} A &= 1.63/1.485 \\ &= 1.098; \quad 1/A = 0.911 \end{aligned}$$

and

$$\begin{aligned}\frac{y_2 - mx_2}{y_1 - mx_2} &= \frac{0.011 - 0.001485}{0.1 - 0.001485} \\ &= 0.096\end{aligned}$$

The N_{OG} value from Equation (10.13) is

$$N_{OG} = 6.82$$

The corresponding values for $y_2 = 0.003, 0.05,$ and 0.065 (employing the same procedure above) are 7.45, 9.66, and 13.5, respectively. ■

ILLUSTRATIVE EXAMPLE 10.8

Determine the packing height of a packed countercurrent absorber required to reduce the Cl_2 concentration in a gas by 99% assuming that a dilute NaOH solution is employed. The following information is given:

Liquid mass flux = $1000 \text{ lb/h} \cdot \text{ft}^2$ (essentially water)

Gas mass flux = $750 \text{ lb/h} \cdot \text{ft}^2$ (essentially air)

Mole fraction of Cl_2 in inlet gas = 0.00043

$H_{OG} = 1.63 \text{ ft}$

SOLUTION: It can be assumed that Cl_2 will react with the dilute NaOH solution. When the solute reacts with the liquid, it can be assumed that $m = 0^{(5)}$ so that Equation (10.15) applies. For a 99% reduction:

$$\begin{aligned}N_{OG} &= \ln(100/1) \\ &= 4.6\end{aligned}$$

The packing height is therefore

$$\begin{aligned}Z &= H_{OG}N_{OG} \\ &= (1.63)(4.6) \\ &= 7.5 \text{ ft}\end{aligned}\tag{10.9}$$

ILLUSTRATIVE EXAMPLE 10.9

Determine the packing height (in feet) of a countercurrent scrubber required to reduce an inlet ammonia concentration by 90%, given the following information:

Liquid molar flux (water) = $1000 \text{ lbmol/h} \cdot \text{ft}^2$

Gas molar flux (air) = $700 \text{ lbmol/h} \cdot \text{ft}^2$

Mole fraction of NH_3 in inlet gas = 0.023

Mole fraction of NH_3 in outlet liquid = 0.015

Assume no NH_3 in inlet water stream

Slope of equilibrium line = 0.93

$H_{OG} = 1.5$ ft

SOLUTION: First check whether the material balance is satisfied:

$$1000(0.015 - 0) = 700(0.023 - 0.0023); \quad y_{\text{out}} = (0.023)(1 - 0.9) \\ 15 = 14.49$$

The material balance is satisfied. Now apply Coburn's equation to calculate N_{OG} :

$$N_{OG} = \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{1 - \frac{1}{A}} \quad (10.13)$$

$$A = \frac{L_m}{mG_m} \\ = 1000 / (0.93)(700) \\ = 1.536$$

Substituting, one obtains

$$N_{OG} = \frac{\ln \left[\left(\frac{0.015 - 0}{0.0015 - 0} \right) \left(1 - \frac{1}{1.536} \right) + \frac{1}{1.536} \right]}{1 - \frac{1}{1.536}} \\ = 4.07$$

Therefore

$$Z = H_{OG}N_{OG}; \quad H_{OG} = 1.5 \text{ ft} \quad (10.9) \\ = (1.5)(4.07) \\ = 6.12 \text{ ft} \quad \blacksquare$$

ILLUSTRATIVE EXAMPLE 10.10

Doyle Unlimited, a Daniel F. Rodenci Corporation, has submitted design plans to Theodore Consultants for a packed ammonia scrubber on an air stream containing NH_3 . The operating and design data provided by Doyle Unlimited, Inc. are given below. Theodore Consultants remember reviewing acceptable plans for a nearly identical scrubber for Doyle Unlimited, Inc. in 2005. After consulting old files, the consultants find all the conditions are identical except for the gas flow rate. What recommendation should be made?

Tower diameter = 3.57 ft

Packed height of column = 8 ft

Gas and liquid temperature = 75°F inlet

Operating pressure = 1.0 atm

Ammonia-free liquid mass flux = 1000 lb/h · ft²

Gas flow rate = 1575 acfm

Inlet NH₃ gas concentration = 2.0 mol%

Air density = 0.0743 lb/ft³

Molecular weight of air = 29

Molecular weight of water = 18

Henry's law constant $m = 0.972$

Figure 10.13; packing type A is used

Colburn chart (Fig. 10.12) applies

Emission regulation = 0.1% NH₃ (by mole or volume)

Packing height = $H_{OG}N_{OG}$

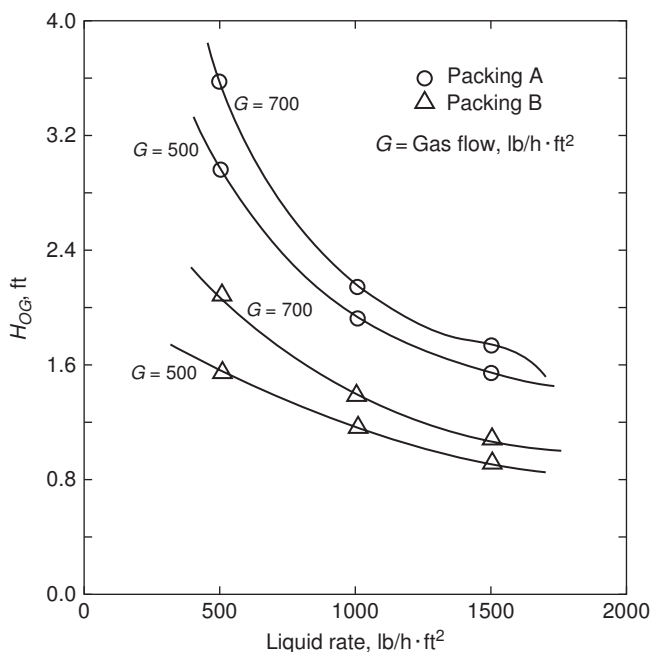


Figure 10.13 H_{OG} values for different types of packing.

SOLUTION: Calculate the cross-sectional area of the tower S (note that S is now employed for the area) in ft²:

$$S = \pi D^2/4 \quad (10.6)$$

Substituting,

$$\begin{aligned} S &= (\pi)(3.57)^2/(4) \\ &= 10.0 \text{ ft}^2 \end{aligned}$$

Calculate the gas molar flux (molar flow rate per unit cross section) and liquid molar flux in lbmol/(ft² · h):

$$\begin{aligned}
 G_m &= q\rho/S(MW)_G \\
 &= (1575)(0.0743)/[(10.0)(29)] \\
 &= 0.404 \text{ lbmol}/(\text{ft}^2 \cdot \text{min}) \\
 &= 24.2 \text{ lbmol}/(\text{ft}^2 \cdot \text{h}) \\
 L_m &= L/(MW)_L \\
 &= (1000)/(18) \\
 &= 55.6 \text{ lbmol}/(\text{ft}^2 \cdot \text{h})
 \end{aligned}$$

The value mG_m/L_m is therefore

$$\begin{aligned}
 mG_m/L_m &= (0.972)(24.2/55.6) \\
 &= 0.423
 \end{aligned}$$

The absorption factor A is defined as

$$A = \frac{L_m}{mG_m} \quad (10.14)$$

Substituting,

$$\begin{aligned}
 A &= \frac{1}{0.423} \\
 &= 2.364
 \end{aligned}$$

The value of $(y_1 - mx_2)/(y_2 - mx_2)$ is

$$\begin{aligned}
 \frac{y_1 - mx_2}{y_2 - mx_2} &= \frac{0.02 - (0.972)(0)}{0.001 - (0.972)(0)} \\
 &= 20.0
 \end{aligned}$$

N_{OG} is calculated from Colburn's equation chart,

$$N_{OG} = \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{1 - \frac{1}{A}} \quad (10.13)$$

Substituting,

$$\begin{aligned}
 N_{OG} &= \frac{\ln \left[(20.0) \left(1 - \frac{1}{2.364} \right) + \left(\frac{1}{2.364} \right) \right]}{1 - \frac{1}{2.364}} \\
 &= 4.30
 \end{aligned}$$

To calculate the height of an overall gas transfer unit, H_{OG} , first calculate the gas mass flux G in $\text{lb}/\text{h} \cdot \text{ft}^2$:

$$\begin{aligned} G &= q\rho/S \\ &= (1575)(0.0743)/10.0 \\ &= 11.7 \text{ lb}/(\text{min} \cdot \text{ft}^2) \\ &= 702 \text{ lb}/(\text{h} \cdot \text{ft}^2) \end{aligned}$$

From Figure 10.13, one obtains (for packing A)

$$H_{OG} = 2.2 \text{ ft}$$

The required packed column height Z , in feet, is

$$\begin{aligned} Z &= N_{OG}H_{OG} \\ &= (4.3)(2.2) \\ &= 9.46 \text{ ft} \end{aligned} \tag{10.9}$$

The proposal should be rejected. ■

ILLUSTRATIVE EXAMPLE 10.11

The calculations for an absorber indicate that it will be excessively tall. Thus, three schemes using two shorter absorbers are considered, as shown on the left hand side of Figure 10.14. Make freehand sketches of operating lines, one for each scheme, showing the relation between the operating lines for the two absorbers and the equilibrium curve. Mark the mole fractions on an equilibrium line–operating line diagram. No calculation is required. Assume dilute solutions. It is suggested that the reader attempt to solve this application prior to looking at the solution which appears in Figure 10.14.

SOLUTION: The solution is presented in the right hand side of Figure 10.14. ■

ILLUSTRATIVE EXAMPLE 10.12

Qualitatively outline how one can size (diameter, height) a packed tower to achieve a given degree of separation without any information on the physical and chemical properties of a gas to be absorbed.

SOLUTION: To calculate the height, one needs both the height of a gas transfer unit H_{OG} and the number of gas transfer units N_{OG} . Since equilibrium data are not available, assume that m (slope of equilibrium curve) approaches zero. This is not an unreasonable assumption for most solvents that preferentially absorb (or react with) the solute. For this condition:

$$N_{OG} = \ln\left(\frac{y_1}{y_2}\right) \tag{10.15}$$

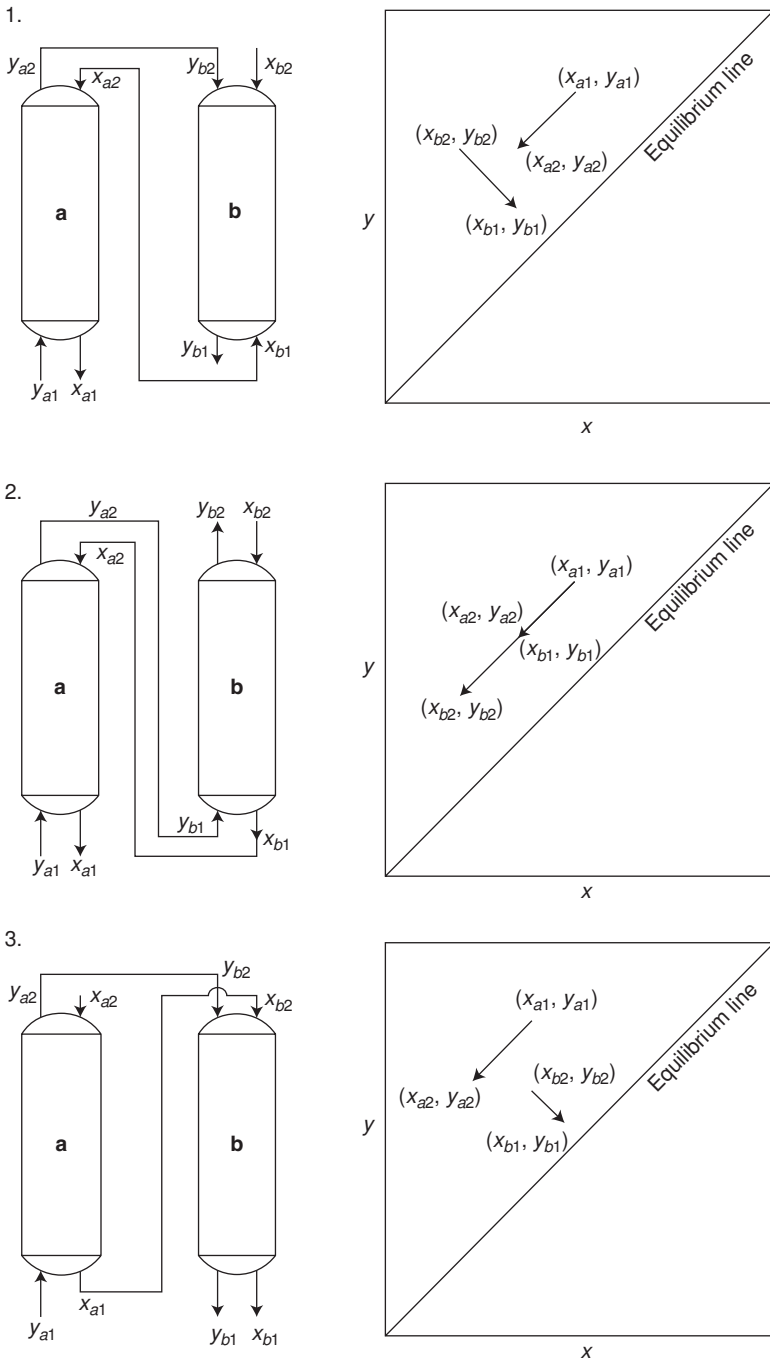


Figure 10.14 Solution to Illustrative Example 10.11.

Table 10.5 Packing Diameter versus H_{OG}

Packing diameter, inches	Plastic packing H_{OG} , feet	Ceramic packing H_{OG} , feet
1.0	1.0	2.0
1.5	1.25	2.5
2.0	1.5	3.0
3.0	2.25	4.5
3.5	2.75	5.5

where y_1 and y_2 represent inlet and outlet concentrations, respectively. Since it is also reasonable to assume the scrubbing medium to be water or a solvent that effectively has the physical and chemical properties of water, H_{OG} can be assigned values usually encountered for water systems. These are given in Table 10.5. For plastic packing, the liquid and gas flow fluxes are both typically in the range of 1500–2000 lb/(h · ft² of cross-sectional area). For ceramic packing, the range of flow rates is 500–1000 lb/h · ft². For difficult-to-absorb gases, the gas flow rate is usually lower and the liquid flow rate higher. Superficial gas velocities (velocity of the gas if the column is empty) are in the 3–6-ft/s range. The height Z is then calculated from

$$Z = (H_{OG})(N_{OG})(SF)$$

where SF is a safety factor, the value of which can range from 1.25–1.5. Pressure drops can vary from 0.15–0.40 inch H₂O/ft packing. Packing size increases with increasing tower diameter. Packing diameters of 1 inch are recommended for tower diameter in the 3 ft range. One should use large packing for larger diameter packing; for smaller towers, smaller packing is usually employed. The reader is left the exercise of verifying the chart in Table 10.6 for plastic packing.⁽⁵⁾ (Note: This problem and design procedure were originally developed by one of the authors in 1985 and later published in 1988.) ■

Table 10.6 Packing Height, Z (ft), as a Function of Efficiency and Plastic Packing Size

Removal efficiency, %	Plastic packing size, inches				
	1.0	1.5	2.0	3.0	3.5
63.2	1.0	1.25	1.5	2.25	2.75
77.7	1.5	1.9	2.25	3.4	4.1
86.5	2.0	2.5	3.0	4.5	5.5
90	2.3	3.0	3.45	5.25	6.25
95	3.0	3.75	4.5	6.75	8.2
98	3.9	4.9	5.9	8.8	10.75
99	4.6	5.75	6.9	10.4	12.7
99.5	5.3	6.6	8.0	11.9	14.6
99.9	6.9	8.6	10.4	15.5	19.0
99.99	9.2	11.5	13.8	20.7	25.3

ILLUSTRATIVE EXAMPLE 10.13

A 1600 acfs gas stream is to be treated in a packed tower containing ceramic packing. The gas stream contains 100 ppm of a solute that is to be reduced to 1 ppm. Estimate the tower's cross-sectional area, diameter, height, pressure drop, and packing size. Use the procedure outlined in Illustrative Example 10.12.

SOLUTION: Key calculations from Illustrative Example 10.12 are provided in Table 10.7 for ceramic packing.⁽⁵⁾ The equation for the cross-sectional area of the tower S in terms of the gas volumetric flow rate q in acfs is (assuming a 4 ft/a superficial velocity)

$$S [\text{ft}^2] = q [\text{acfs}]/4.0$$

An equation to estimate the tower packing pressure drop ΔP in terms of Z is

$$\Delta P [\text{in H}_2\text{O}] = (0.2)Z; \quad Z = \text{ft}$$

The following packing size(s) is (are) recommended:

For $D \approx 3$ ft, use 1-inch packing

For $D < 3$ ft, use <1-inch packing

For $D > 3$ ft, use >1-inch packing

As noted earlier, recommended packing size increases with increasing diameter.

For the problem at hand

$$S = 1600/4 = 400 \text{ ft}^2$$

The diameter D is

$$D = (4S/\pi)^{0.5} \quad (10.6)$$

Table 10.7 Packing Height, Z (ft), as a Function of Efficiency and Ceramic Packing Size

Removal efficiency, %	Ceramic packing size, inches				
	1.0	1.5	2	3	3.5
63.2	2.0	2.5	3.0	4.5	5.5
77.7	3.0	3.7	4.5	6.75	8.25
86.5	4.0	5.0	6.0	9.0	11.0
90	4.6	5.75	6.9	10.4	12.7
95	6.0	7.5	9.0	13.5	16.5
98	7.8	9.8	11.7	17.6	21.5
99	9.2	11.5	13.8	20.7	25.3
99.5	10.6	13.25	15.9	23.8	29.1
99.9	13.8	17.25	20.7	31.1	38.0
99.99	18.4	23.0	27.6	41.4	50.7

Substituting,

$$\begin{aligned} D &= [(4)(400)/\pi]^{0.5} \\ &= 22.6 \text{ ft} \end{aligned}$$

For a tower diameter this large, the 3.5-inch packing should be used. Additional information is available in the literature.⁽⁵⁾ ■

Pressure Drop

The pressure drop through a packed column for any combination of liquid and gas flows in the operable range is an important economic consideration in the design of such columns. For most random packings, the pressure drop incurred by the gas is influenced by the gas and liquid flow rates. At constant gas rate, an increase in liquid throughput—which takes up more room in the packing (increased holdup) and, therefore, leaves less room for the gas (greater restriction)—is accompanied by an increase in pressure drop until the liquid flooding rate is reached. At this point, any slight liquid excess that cannot pass through remains atop the packing, building up a deeper and deeper head (or pressure drop), hypothetically reaching an infinite value. Similarly, at constant liquid downflow, increasing the gas flow is again accompanied by a rising pressure drop until the flooding rate is reached, whereupon the slightest gas increase will cause a decline in permissible liquid throughput. This causes the liquid to again accumulate atop the packing, so that pressure drop again continues to increase. For a particular packing, the most accurate pressure drop data will be those available directly from the manufacturer. However, for the purposes of estimation, Figure 10.11 is simple to use and usually provides reasonable results.

Some general “rules of thumb” in the design of packed columns do exist. They are by no means final in that there are other considerations that might have to be taken into account (allowable pressure drop, possible column height restrictions, and so on). The rules must therefore be applied discriminately. For approximation purposes, if the gas rate is greater than about 500 acfm, a nominal packing size smaller than 1 inch would probably not be practical: similarly, at about 2000 acfm, sizes smaller than 2 inches would also likely be impractical. The nominal size of the packing should not exceed about 1/20th of the column diameter. The usual practice is to design so that the operating gas rate is approximately 75% of the rate that would cause flooding. If possible, column dimensions should be in readily available sizes (i.e., diameters to the nearest half foot and heights to the nearest foot). If the column can be purchased “off-the-shelf”, as opposed to being specially made, substantial savings can be realized.

ILLUSTRATIVE EXAMPLE 10.14

A packed column is designed to absorb ammonia from a gas stream. The unit operates at 60% of the flooding gas mass velocity, the actual liquid flow rate is 25% more than the minimum, and 90% ammonia to be collected. Given the operating conditions and type of packing below, calculate the height of packing, the column diameter, and the operating pressure drop. Data

includes: gas mass flow rate = 5000 lb/h; NH_3 concentration in inlet gas stream = 2.0 mol%; scrubbing liquid is pure water; packing type = 1 inch Raschig rings; H_{OG} of the column = 2.5 ft; Henry's law constant, $m = 1.20$; density of gas (air) = 0.075 lb/ft^3 ; density of water = 62.4 lb/ft^3 ; viscosity of water = 1.8 cP, $\psi = 1$ (ratio of liquid to water density); and $F = 160$ (for 1 inch Raschig rings).

SOLUTION: First calculate the equilibrium outlet liquid composition, and the outlet gas composition for 90% removal:

$$x_1^* = \frac{y_1}{m} \quad (10.1)$$

Substituting,

$$x_1^* = \frac{0.02}{1.20} = 0.0167$$

The minimum liquid to gas ratio (molar basis) is obtained by a material balance:

$$y_2 = \frac{0.1y_1}{[(1 - y_1) + 0.1y_1]} = \frac{0.1(0.02)}{[(1 - 0.02) + 0.1(0.02)]} = 0.00204$$

The minimum liquid–gas ratio (molar basis) is obtained by a material balance:

$$\left(\frac{L_m}{G_m}\right)_{\min} = \frac{y_1 - y_2}{x_1^* - x_2} \quad (10.4)$$

Substituting,

$$\left(\frac{L_m}{G_m}\right)_{\min} = \frac{0.02 - 0.00204}{0.0167 - 0} = 1.08$$

The actual ratio is 25% above the minimum. Thus,

$$\left(\frac{L_m}{G_m}\right)_{\text{act}} = 1.25 \left(\frac{L_m}{G_m}\right)_{\min} = 1.25(1.08) = 1.35$$

Two parameters are needed to use the Colburn chart to evaluate N_{OG} :

$$\frac{y_1 - mx_2}{y_2 - mx_2} = \frac{0.02 - 1.2(0)}{0.00204 - 1.2(0)} = 9.80$$

and

$$\frac{mG_m}{L_m} = \frac{1.2}{1.35} = 0.889$$

From Colburn's Chart (Fig. 10.12)

$$N_{OG} = 6.2$$

The packing height is then

$$Z = N_{OG}H_{OG} \quad (10.9)$$

Substituting,

$$\begin{aligned} Z &= 6.2(2.5) \\ &= 15.5 \text{ ft} \end{aligned}$$

Figure 10.11 is employed to calculate the tower diameter and packing pressure drop,

$$\begin{aligned}\frac{L}{G} \left(\frac{\rho}{\rho_L} \right)^{0.5} &= 1.35 \left(\frac{18}{29} \right) \left(\frac{0.075}{62.4} \right)^{0.5} \\ &= 0.0291\end{aligned}$$

From Figure 10.11,

$$\frac{G^2 F \Psi \mu_L^{0.2}}{\rho_L \rho g_c} = 0.19$$

Thus, the flooding mass velocity is

$$\begin{aligned}G_f &= \sqrt{\frac{0.19 \rho_L \rho g_c}{F \Psi \mu_L^{0.2}}} = \sqrt{\frac{0.19(62.4)(0.075)(32.2)}{160(1)(1.8)^{0.2}}} \\ &= 0.40 \text{ lb/ft}^2 \cdot \text{s}\end{aligned}$$

The actual velocity is

$$\begin{aligned}G &= 0.6G_f = 0.6(0.40) \\ &= 0.24 \text{ lb/ft}^2 \cdot \text{s} = 864 \text{ lb/ft}^2 \cdot \text{h}\end{aligned}$$

The tower diameter may now be calculated directly from

$$D = 1.13S^{0.5} \tag{10.6}$$

Substituting

$$\begin{aligned}D &= 1.13 \left(\frac{W}{fG} \right)^{0.5} = 1.13 \left(\frac{5000}{(1)(864)} \right)^{0.5} \\ &= 2.72 \text{ ft}\end{aligned}$$

The operating pressure drop can be estimated from Figure 10.11. At 60% of flooding, the ordinate becomes

$$(0.19)(0.6)^2 = 0.068$$

Employing this ordinate and an abscissa of 0.029 gives an operating pressure drop of approximately

$$\Delta P = 0.065 \text{ in H}_2\text{O/ft packing}$$

Calculating the overall pressure drop across the column is left as an exercise for the reader. ■

ILLUSTRATIVE EXAMPLE 10.15

Consider the absorber system shown in Figure 10.15. Corenza Engineers designed the unit to operate with a maximum discharge concentration of 50 ppm. Once the unit was installed and running, the unit operated with a discharge of 60 ppm. Rather than purchase a new unit, what options are available to bring the unit into compliance with the specified design concentration?

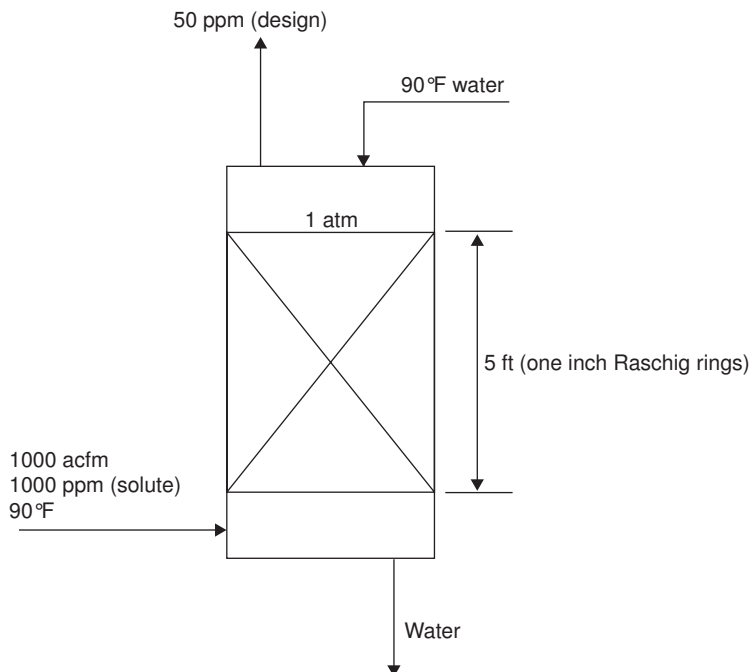


Figure 10.15 Absorber failure to meet design performance.

SOLUTION: This is obviously an open-ended question. On the basis of the material presented earlier and the solutions to several of the problems in this chapter, one may employ any one or a combination of several suggestions recommended by Theodore.⁽⁶⁾ The reader is referred to Illustrative Example 20.4 in Chapter 20 for the “solution” to this illustrative example. ■

DESIGN AND PERFORMANCE EQUATIONS—PLATE COLUMNS

The most important design considerations for plate columns include the calculation of the column diameter, type and number of plates to be used (usually either bubble-cap or sieve plates), actual plate layout and physical design, and plate spacing; these, in turn, determine column height. To consider each of these to any great extent is beyond the scope of this chapter, particularly since it received attention in Chapter 9. The discussion that follows, therefore, will be a relatively concise presentation of some of the general absorber design techniques that will provide satisfactory results for purposes of estimation.⁽⁷⁻⁹⁾

The column diameter, and consequently its cross section, must be sufficiently large to handle the gas and liquid at velocities that will not cause flooding or excessive entrainment. The superficial gas velocity for a given type of plate at flooding is given

by the relation

$$V_F = C_F \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{0.5} \quad (10.16)$$

where V_F (the notation usually employed for plate columns) is the gas velocity through the net column cross sectional area for gas flow, $\text{ft}^3/\text{s} \cdot \text{ft}^2$, the densities are in lb/ft^3 , and C_F is an empirical coefficient that depends on the type of plate and operating conditions.

The net cross section is the difference between the column cross section and the area taken up by downcomers. In actual design, some percent of V_F is usually used—for nonfoaming liquids 80–85% of V_F and 75% or less for foaming liquids. Of course, the value is subject to a check of entrainment and pressure drop characteristics. The calculation of column diameter based on Equation (10.16) assumes that the gas flow rate is the controlling factor in its determination.

After a plate layout has been assumed, it is then necessary to check the plate for its liquid handling capacity. If the liquid-to-gas ratio is high and the column diameter large, the check will indicate whether the column will show a tendency toward flooding or gas maldistribution on the plate. If this is the case, then the liquid rate is the controlling factor in estimating the column diameter and a satisfactory assumption for design purposes is a plate-handling capacity of 30 gal/min of liquid per foot of diameter.⁽⁹⁾ However, well-designed single-pass cross-flow plate can ordinary be expected to handle up to 60 gal/min of liquid per foot of diameter without an excessive liquid gradient. It should also be noted that low gas rates can lead to *weeping*, a condition where the liquid flows down through the holes in the plate rather than across the plate.

The column height is determined from the product of the number of actual plates (theoretical plates divided by the overall plate efficiency) and the plate spacing chosen. The theoretical plate (or stage) is the theoretical unit of separation in plate column calculations. It is defined as a plate in which two dissimilar phases are brought into intimate contact with each other and then are physically separated. During the contact, various diffusing components of the mixture redistribute themselves between the phases. In an equilibrium stage, the two phases are well mixed for a sufficient time to allow establishment of equilibrium between the phases leaving the stage. In effect, no further net change of composition of the phases is possible at equilibrium for a given set of operating conditions. The number of theoretical plates can be determined graphically from the operating diagram composed of an operating line and equilibrium curve.

In the above discussion of equilibrium stages, it was assumed that the phases leaving the stage were in equilibrium. In actual countercurrent multistage equipment, it is not practical to provide the combination of residence time and intimacy of contact required to accomplish equilibrium. Hence, the concentration change for a given stage is less than that predicted by equilibrium considerations. Stage efficiencies are employed to characterize this condition. The efficiency term frequently used is the overall stage (plate efficiency), given by the ratio of theoretical contacts required for a given separation to the actual number of contacts required for the same operation.

While reliable information on such an efficiency is most desirable and convenient to use, so many variables come into play that really reliable values for the overall stage efficiency are difficult to come by. This value is generally obtained by experiment or field test data, or may be specified by the vendor.

The number of theoretical plates may be determined directly without recourse to graphical techniques for cases where both the operating line and the equilibrium curve may be considered straight (dilute solutions). This will frequently be the case for a relatively dilute gas (as usually encountered in air pollution control) and liquid solutions where, more often than not, Henry's law is usually applicable. Since the quantity of gas absorbed is small, the total flows of liquid and gas entering and leaving the column again essentially remain constant. Hence, the operating line will be substantially straight. For such cases, the Kremser–Brown–Sounders^(10,11) equation applies for determining the number of theoretical plates, N_p :

$$N_p = \log \frac{\left[\left(\frac{y_{Np+1} - mx_0}{y_1 - mx_0} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{\log A} \quad (10.17)$$

Note that \ln may be employed rather than \log in both the numerator and denominator. Here mx_0 is the gas composition in equilibrium with the entering liquid (m is Henry's law constant = slope of the equilibrium curve). If the entering liquid contains no solute gas, then $x_0 = 0$ and Equation (10.17) can be simplified further. The solute concentrations in the gas stream, y_{Np+1} and y_1 represent inlet and outlet conditions, and L and V (that appear in A) the total mole rates of liquid and gas flow per unit time per unit column cross-sectional area. Small variations in L and V may be roughly compensated for by using the geometric average value of each taken at the top and bottom of the column. Equation (10.17) has been plotted in Figure 10.16 for convenience and may be used for the solution to this equation.

Chen⁽¹²⁾ derived a simplified algebraic equation that could be used to estimate the theoretical plates, n , in either an absorber or stripper. The final equation took the form (retaining Chen's notation):

$$A^n = \frac{y_b + \phi}{y_n + \phi} \quad (10.18)$$

where

$$\phi = \frac{y_t - A(B + mx_t)}{A - 1} \quad (10.19)$$

and A is the absorption factor, y_t is the top plate gas mole fraction, y_b is the bottom plate gas mole fraction and $y_n = B + mx_t$ (equilibrium line) is the gas mole fraction at plate n .

The number of actual trays, which is based on the tray efficiency, is determined by the mechanical design and conditions of operation. For the case where the equilibrium curve and operating lines are straight, the overall tray efficiency E_0 can be computed

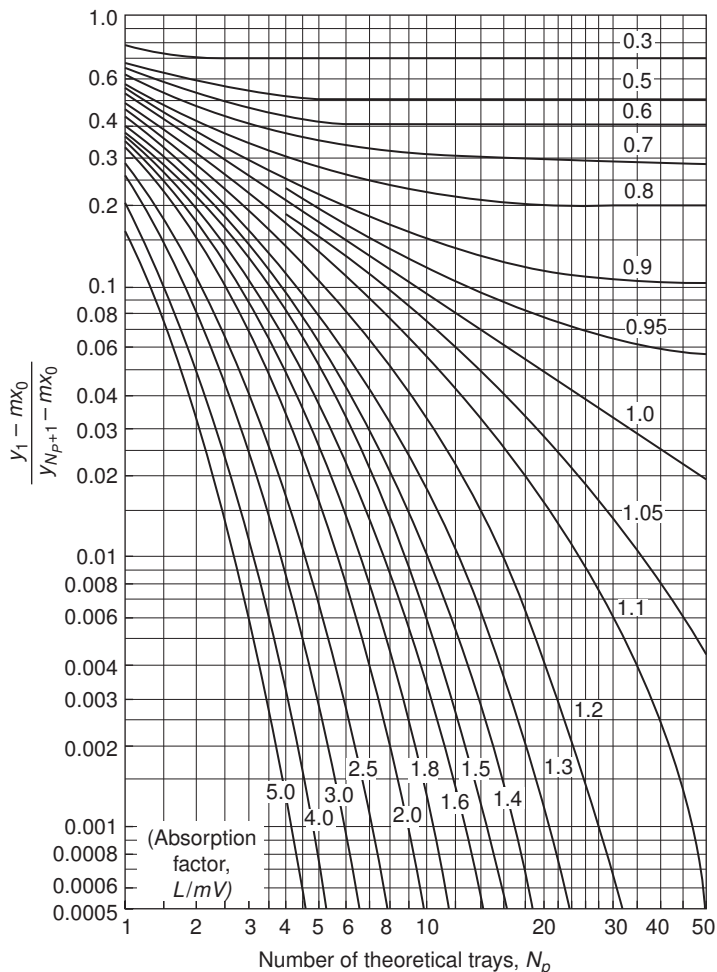


Figure 10.16 Number of theoretical stages for countercurrent absorption columns.

and the number of actual trays determined analytically:

$$\begin{aligned}
 E_0 &= \frac{\text{equilibrium trays}}{\text{actual trays}} \\
 &= \frac{\log(1 + E_{\text{MGE}})(1/A - 1)}{\log(1/A)} \quad (10.20)
 \end{aligned}$$

where E_{MGE} = Murphree efficiency, as noted in Chapter 9, corrected for entrainment (values available in the literature). Empirical data for standard tray designs within standard ranges of liquid and gas rates are available. These data, as shown in Figure 10.17, are accurate for bubble-cap trays and can be used as rough estimates for sieve and valve

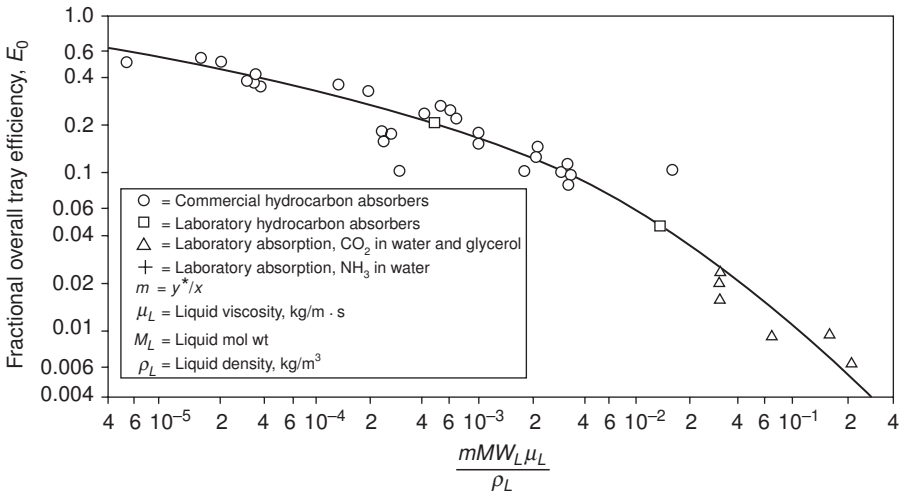


Figure 10.17 Overall tray efficiencies of bubble-cap tray absorbers.

trays. After the overall efficiency of the tower is determined, the number of actual trays is calculated using:

$$N_{\text{act}} = \frac{N}{E_0} = \frac{N_p}{E_0} \quad (10.21)$$

The general procedure to follow in sizing a plate tower is given below.⁽¹³⁾

- 1 Calculate the number of theoretical stages, N , using Figure 10.16 or Equation (10.17).
- 2 Estimate the efficiency of separation, E . This may be determined at the local (across plate), plate (between plates), or overall (across column) level. The overall efficiency, E_0 , is generally employed.
- 3 Calculate the actual number of plates:

$$N_{\text{act}} = \frac{N}{E_0} \quad (10.21)$$

- 4 Obtain the height between plates, h . This is usually in the 12- to 36-inch range. Many towers use a 24-inch plate spacing.
- 5 The tower height, Z , is then

$$Z = N_{\text{act}} h \quad (10.22)$$

- 6 The diameter may be calculated directly from Equation (10.16).

- 7 The plate or overall pressure drop is difficult to quantify accurately. It is usually in the 2- to 6-inch H₂O per plate range for most columns with the lower and upper values applying to small and large diameters, respectively.

ILLUSTRATIVE EXAMPLE 10.16

Refer to Illustrative Example 10.7. Repeat the calculations as they apply to a plate column. In effect, determine the number of theoretical plates N . Employ Equation (10.17).

SOLUTION: Equation (10.17) applies for a plate column:

$$N = \frac{\log \left[\left(\frac{y_{N+1} - mx_0}{y_1 - mx_0} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{\log A} \quad (10.17)$$

once again

$$m = 1.485$$

$$(L_m/G_m)_{\text{act}} = 1.2(L_m/G_m)_{\text{min}}$$

Results for N_p are provided in Table 10.8. ■

Table 10.8 Number of Plates for Illustrative Example 10.16

X_2	$(L_m/G_m)_{\text{min}}$	$(L_m/G_m)_{\text{act}}$	A	N_p
0	1.384	1.661	1.119	5.48
0.001	1.4031	1.6837	1.1339	5.87
0.003	1.4438	1.7326	1.1667	6.93
0.005	1.4872	1.7846	1.2018	8.82
0.0065	1.5215	1.8258	1.2295	12.21

ILLUSTRATIVE EXAMPLE 10.17

In an attempt to quantify the effect of enthalpy of solution effects on the absorption of HCl into scrubbing water in an absorber, Pallechi Consultants reviewed the literature⁽⁶⁾ and obtained the following rough estimates of this effect. The data provided temperature increases as a function of increasing HCl concentration (mass percent basis) in water:

$$0-1.5\% = 10^\circ\text{C}$$

$$0-3.0\% = 15^\circ\text{C}$$

$$0-5.0\% = 20^\circ\text{C}$$

Apply the above data and estimate the discharge temperature increase for the following two HCl scenarios:

Scenario 1: 0.0% inlet to 1.5% outlet (mass percent)

Scenario 2: 0.5% inlet to 3.0% outlet (mass percent)

SOLUTION: Since enthalpy is a point function, it is reasonable to assume that the temperature effects are additive. Therefore, the temperature increases are

Scenario 1:

$$\begin{aligned}\Delta T &= \Delta T_{1.5} - \Delta T_{0.0} \\ &= 10 - 0 \\ &= 10^\circ\text{C}\end{aligned}$$

Scenario 2:

$$\begin{aligned}\Delta T &= \Delta T_{3.0} - \Delta T_{1.5} \\ &= 15 - 10 \\ &= 5^\circ\text{C}\end{aligned}$$

The reader should note that this is an effect that often should be reflected in engineering applications since any increase in the temperature of the scrubbing liquid adversely affects the equilibrium, reducing the equilibrium capacity of the liquid. Note that it is the temperature of the liquid, *not* the gas, that affects equilibrium. ■

ILLUSTRATIVE EXAMPLE 10.18

You are requested to outline a procedure to calculate the height of a plate tower required to absorb $R\%$ of organics (A and B) in an organic–air mixture. The following assumptions can be made:

- 1 Ideal gas and liquid solutions
- 2 Isothermal operation
- 3 Liquid-to-gas molar flow rate ratio (L_m/G_m) through the tower is constant
- 4 Henry's law applies; dimensionless constants available
- 5 Plate efficiency, E_0 , is given
- 6 Height between stages, h , is also given
- 7 Absorbing liquid contains no A and B
- 8 Inlet gas concentrations of A and B are given

Calculate the height of the tower for a methylamine (A)/dimethylamine–(B) air–water system if:

$$L_m/G_m = 1.0$$

The absorption factors are, $A_i = L_m/m_iG_m$; $A_A = 0.85$, $A_B = 0.75$. In addition,

$$\begin{aligned}E_0 &= 0.5 \\h &= 2.0 \text{ ft} \\y_{A,i} &= 0.01 \\y_{B,i} &= 0.008\end{aligned}$$

The required recovery efficiency of the tower, $R = 77.78\%$ (total).

SOLUTION: The Kremser–Souter–Brown (KSB) equation, referred to by some as the Kremser equation, applies individually to both components A and B , and may be written in the following form:

$$\frac{y_{N+1} - y_1}{y_{N+1} - mx_0} = \frac{A^{N+1} - A}{A^{N+1} - 1} \quad (10.23)$$

where A is the absorption factor, y_{N+1} is the inlet gas concentration, y_1 is the outlet concentration, x_0 is the inlet absorbing liquid concentration, and N is the number of theoretical plates required. Generally, one also assumes no condensation, no mixing (heat) effects, no chemical reaction, etc. Since the concentration is usually dilute, the liquid and gas rates are also assumed to be constant. Thus, the absorption factor A is also constant. The key to the multicomponent calculation suggested here is to assume:

- 1 no interaction effects between the various components, and
- 2 the absorption of each component occurs as if the other components are not present (i.e., treat each component separately); the KSB equation is then employed for each species present in the gas mixture. Note also that the assumption of an ideal solution and isothermal operation is valid for many operations.

For component A :

$$\frac{0.01 - y_{A1}}{0.01 - 0} = \frac{0.85^{N+1} - 0.85}{0.85^{N+1} - 1} \quad (1)$$

For component B :

$$\frac{0.008 - y_{B1}}{0.008 - 0} = \frac{0.75^{N+1} - 0.75}{0.75^{N+1} - 1} \quad (2)$$

Calculate the total outlet concentration of the gas mixture:

$$y_{A,1} + y_{B,1} = (1 - R)(y_{A,N+1} + y_{B,N+1}) = (1 - 0.7778)(0.01 + 0.008) = 0.0040 \quad (3)$$

Solve the three equations obtained above for $N + 1$, $y_{A,1}$, and $y_{B,1}$. The result via a trial-and-error calculation is

$$\begin{aligned}N + 1 &= 10 \\y_{A,1} &= 0.00187 \\y_{B,1} &= 0.00212\end{aligned}$$

Note that a simplified equation is available for calculating the total outlet concentration or loading for a given unit when more than two components are absorbed:

$$\sum_i^n y_{i,1} = \sum_i^n \frac{(1 - A_i)(y_{i,in})}{1 - A_i^{N+1}} \quad (10.24)$$

where i is the component in question and n is the number of components.

Using the above results, calculate the number of theoretical plates required and the height of the plate tower, Z :

$$\begin{aligned} N &= (N + 1) - 1 = 9 \\ Z &= \frac{Nh}{E_0} \\ &= \frac{9(2)}{0.5} = 36 \text{ ft} \quad \blacksquare \end{aligned} \quad (10.21)$$

The reader should note that there presently exists little to no information available in the literature for a simple treatment of multicomponent absorption. The chemical engineering literature does provide a “shortcut” method where variations in flow rates and temperature are taken into account. However, the shortcut method requires a double trial-and-error calculation. A rigorous technique—involving tray to tray calculations—is also available. The method presented here, assuming ideal conditions, can also be solved graphically. Furthermore, it can be set up to solve for either the outlet concentration or the required liquid flow to achieve a given separation in a particular tower. Finally, this simplified technique can also be used to design or predict the performance of packed towers. Here $Z = (H_{OG})_A(N_{OG})_A = (H_{OG})_B(N_{OG})_B = \dots$ where it has been assumed $(H_{OG})_A = (H_{OG})_B$.

STRIPPING

Quite often, an absorption column is followed by a liquid absorption process in which the gas solute is removed from the absorbing medium by contact with an insoluble gas. This operation is called “stripping” and is utilized to regenerate the solute “rich” solvent so that it can be recycled back to the absorption unit. The rich solution enters the stripping unit and the volatile solute is stripped from solution by either reducing the pressure, increasing the temperature, using a stripping gas to remove the vapor solute dissolved in the solvent, or any combination of these process options. While the concept of stripping is opposite to that of absorption, it is treated in the same manner. The operating line developed for absorption (see Illustrative Example 10.1) can be applied to a stripping unit (see Fig. 10.18 for component A).

As was developed in Equation (10.4), the operating line for a stripping unit is also given by

$$\frac{L_m}{G_m} = \frac{y_{A1} - y_{A2}}{x_{A1} - x_{A2}} \quad (10.4)$$

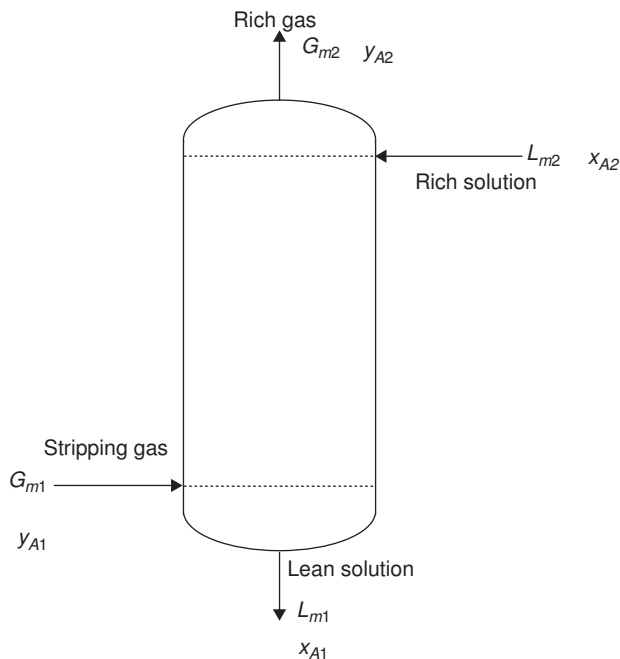


Figure 10.18 Stripping unit.

However, since this process is the opposite of absorption, solute is transferred from the liquid to the gas, and thus the operating line lies below the equilibrium curve. When absorption was addressed, a minimum liquid to gas ratio, $(L_m/G_m)_{\min}$, could be set in order to determine limits on the design. However, for stripping operations, a minimum gas to liquid ratio, $(G_m/L_m)_{\min}$, is now used and corresponds to the minimum gas rate required to achieve the desired separation. The minimum gas to liquid ratio can be found by the following procedure.

- 1 As shown in Figure 10.19, a line is drawn from the point (y_{A1}, x_{A1}) , which represents of the mole fractions of solute in the stripping gas feed and the stripped liquid stream, respectively, to the intersection of $x = x_{A2}$ (which represents the mole fraction of the liquid stream to be stripped) and the equilibrium line. If a plot of the equilibrium data results in a curve, then a tangent is drawn on the curve at the point corresponding to the value of the inlet rich solution concentration (x_{A2}) .
- 2 Obtain y_{A2} by reading or calculating the corresponding value on the y-axis.
- 3 Rearrange Equation (10.5) in terms of $(G_m/L_m)_{\min}$ and insert the known quantities.
- 4 Multiply the results of (3) by the liquid flow rate to obtain the minimum stripping gas rate.
- 5 Typically, an actual value of $1.3(G_m/L_m)_{\min}$ is employed to assure an efficient separation.

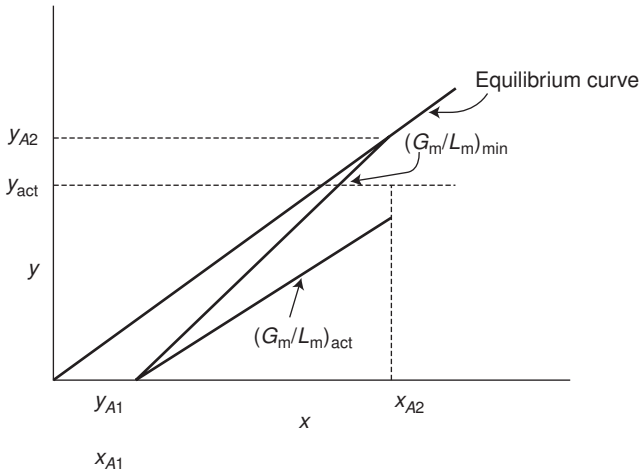


Figure 10.19 Minimum gas to liquid ratio.

Other calculations essentially remain the same for both packed and plate towers, except that the height of a packed tower is given by

$$Z = H_{OL}N_{OL} \quad (10.25)$$

where

$$N_{OL} = \frac{\ln \left[\left\{ \frac{x_2 - (y_1/m)}{x_1 - (y_1/m)} \right\} (1 - A) + A \right]}{1 - A} \quad (10.26)$$

ILLUSTRATIVE EXAMPLE 10.19

Following the absorption of ethylene oxide (EO) from a process stream with water, the water stream is stripped of the EO as part of the regeneration step using steam. For a feed EO concentration of 0.4 mol%, determine the actual amount of steam required for stripping EO to a concentration of 0.02 mol% if the liquid flow rate is 500 lbmol/h and 30% excess EO-free steam is required for the separation in a packed column with 1 inch Raschig rings. Also, determine the diameter of the column. Assuming that Henry's law applies, use the equation $y = 20x$ as the EO equilibrium data for this system. The system is at 30 psia and employs saturated steam at system conditions.

SOLUTION: Generate the EO equilibrium data for $x = 0$ to 0.008 if $y = 20x$ (see Table 10.9). The plot of the equilibrium line is provided in Figure 10.20.

Determine the minimum gas to liquid ratio. Refer to Figure 10.20.

$$y_{A2} = 0.08$$

Table 10.9 Ethylene-Oxide Equilibrium Data

y	x
0.01	0.0005
0.02	0.001
0.03	0.0015
0.04	0.002
0.05	0.0025
0.06	0.003
0.07	0.0035
0.08	0.004
0.09	0.0045
0.10	0.005
0.11	0.0055
0.12	0.006

Since

$$\frac{G_m}{L_m} = \frac{x_{A1} - x_{A2}}{y_{A1} - y_{A2}} \tag{10.5}$$

Substituting gives

$$\begin{aligned} \left(\frac{G_m}{L_m}\right)_{\min} &= \frac{0.0005 - 0.004}{0 - 0.08} \\ &= 0.04375 \end{aligned}$$

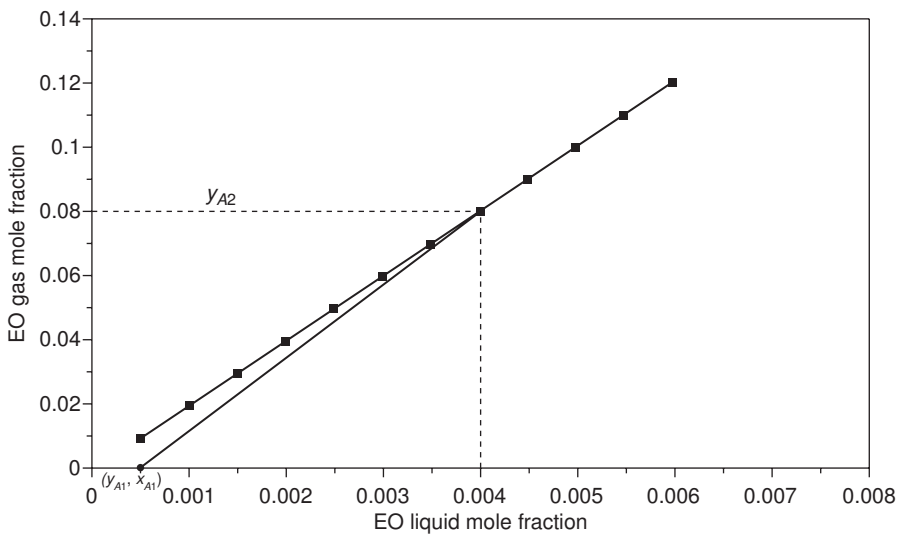


Figure 10.20 Stripping of EO-water system.

The actual steam rate is therefore

$$\begin{aligned}\left(\frac{G_m}{L_m}\right)_{\text{act}} &= (1.3)0.04375 \\ &= 0.0569\end{aligned}$$

and

$$\begin{aligned}\dot{m}_{\text{st}} &= 0.0569(500)(18) \\ &= 511.8 \text{ lb steam/h}\end{aligned}$$

Determine the abscissa, $\frac{L}{G} \left(\frac{\rho_G}{\rho_L}\right)^{0.5}$ from Figure 10.11:

$$\frac{L}{G} \left(\frac{\rho_G}{\rho_L}\right)^{0.5} = \left(\frac{L_m}{G_m}\right) \left(\frac{18}{18}\right) \left(\frac{\rho_G}{\rho_L}\right)^{0.5}$$

The density of steam can be found from the ideal gas law:

$$\begin{aligned}\rho &= \frac{P(MW)}{RT} = \frac{(30)(18)}{(10.73)(250.34 + 460)} \\ &= 0.07085 \text{ lb/ft}^3\end{aligned}$$

Note that the temperature used is the saturated steam condition at 30 psia. Since $\rho_L = 62.4 \text{ lb/ft}^3$,

$$\frac{1}{0.0569} \frac{18}{18} \left(\frac{0.07085}{62.4}\right)^{0.5} = 0.592$$

From Figure 10.11, the ordinate is 0.048.

Solve for the flooding gas mass velocity, G_f , in $\text{lb/ft}^2 \cdot \text{s}$.

$$0.048 = \frac{G_f^2 F \Psi \mu_L^{0.2}}{\rho_L \rho g_c}$$

Substituting

$$\Psi = 1 \text{ (ratio of water to liquid density)}$$

$$\mu_L = 0.19 \text{ cP}$$

$$F = 160 \text{ (for 1 inch Raschig rings)}$$

$$g_c = 32.2 \text{ ft} \cdot \text{lb/lb}_f \cdot \text{s}^2$$

leads to

$$G_f = 0.0595 \text{ lb/s} \cdot \text{ft}^2$$

The actual gas mass velocity, G_{act} , is 60% of the flooding velocity:

$$G_{\text{act}} = 0.6G_f = 0.6(0.0595) = 0.0357 \text{ lb/s} \cdot \text{ft}^2$$

Finally, calculate the diameter of the column in ft:

$$\begin{aligned}
 S &= \frac{\dot{m}_{\text{st}}}{G_{\text{act}}} = \frac{511.8}{0.0375(3600)} \\
 &= 3.98 \text{ ft}^2 \\
 D &= 1.13S^{0.5} \\
 &= 2.25 \text{ ft}
 \end{aligned} \tag{10.6}$$

ILLUSTRATIVE EXAMPLE 10.20

An atmospheric packed tower air stripper is used to clean contaminated ground-water with a concentration of 100 ppm trichloroethylene (TCE). The stripper was designed such that the packing height is 13 ft, the diameter is 5 ft, and the height of a transfer unit (HTU) is 3.25 ft. Assume that Henry's law applies with a constant (H) of 324 atm at 68°F. Also, at these conditions, the molar density of water is 3.47 lbmol/ft³ and the air–water mole ratio (G_m/L_m) is related to the air–water volume ratio (G''/L'') through $G''/L'' = 130 G_m/L_m$, where the units of G'' and L'' are ft³/s · ft². If the stripping factor (R) used in the design is 5.0, what is the removal efficiency?

In addition, the following equation has been developed for the calculation of the number of transfer units (NTUs) for an air–water stripping system and is based on the stripping factor R and the inlet/outlet concentrations:

$$\text{NTU} = \left(\frac{R}{R-1} \right) \ln \left(\frac{(C_{\text{in}}/C_{\text{out}})(R-1) + 1}{R} \right) \tag{10.27}$$

where C_{in} = inlet contaminant concentration, ppm

C_{out} = outlet contaminant concentration, ppm

R = stripping factor

SOLUTION: As described earlier, the height of packed tower can be calculated by

$$Z = (N_{OG})(H_{OG}) \cong (\text{NTU})(\text{HTU})$$

Rearranging Equation (10.27), one obtains

$$\begin{aligned}
 C_{\text{out}} &= \frac{C_{\text{in}}(R-1)}{R \exp[(\text{NTU})(R-1)/R] - 1}; \quad \text{NTU} = 13/3.25 \\
 &= \frac{(100)(5.0-1)}{(5.0) \exp[(13/3.25)(5.0-1)/5.0] - 1} \\
 &= 3.3
 \end{aligned}$$

The removal efficiency (RE) is then

$$\begin{aligned}
 \text{RE} &= [(C_{\text{in}} - C_{\text{out}})/C_{\text{in}}]100\% \\
 &= [(100 - 3.3)/100]100\% \\
 &= 96.7\%
 \end{aligned}$$

PACKED VS PLATE TOWER COMPARISON

Of the various types of gas absorption devices, packed columns and plate columns are the most commonly used in practice. Although packed columns are used more often, both have their special areas of usefulness, and the relative advantages and disadvantages of each are worth considering. In general:

- 1 The pressure drop of the gas passing through the packed column is smaller.
- 2 The plate column can stand an arbitrarily low liquid feed and permits a higher gas feed than the packed column. It can also be designed to handle liquid rates that would ordinarily flood the packed column.
- 3 If the liquid deposits a sediment, the plate column is more advisable. By fitting the column with manholes, the plate column can be cleaned of accumulated sediment that would clog many packing materials and warrant necessary costly removal and refilling of the column. Packed columns are also susceptible to plugging if the gas contains particulate contaminate(s).
- 4 In mass transfer processes accompanied by considerable heat effects, cooling or heating the liquid is much easier in the plate column. A system of pipes immersed in the liquid can be placed on the plates between the caps, and heat can be removed or supplied through the pipe wall directly to the area in which the process is taking place. The solution of the same problem for a packed column leads to the division of this process into a number of sections, with the cooling or heating of the liquid taking place between these sections.
- 5 The total weight of the plate column is usually less than the packed column designed for the same capacity.
- 6 A well-installed plate column avoids serious channeling difficulties insuring good, continuous contact between the gas and liquid throughout the column.
- 7 In highly corrosive atmospheres, the packed column is simpler and cheaper to construct.
- 8 The liquid holdup in the packed column is considerably less than in the plate column.
- 9 Temperature changes are apt to do more damage to the packed column than to the plate column.
- 10 Plate columns are advantageous for absorption processes with an accompanying chemical reaction (particularly when it is not very rapid). The process is favored by a long residence time of the liquid in the column and by easier control of the reaction.
- 11 Packed columns are preferred for liquids with high foaming tendencies.
- 12 The relative merits of the plate column and packed column for a specified purpose are normally determined only by comparison of the actual cost figures resulting from a detailed design analysis for each type. Most conditions

being equal, packed columns in the smaller sizes (diameters up to 2 or 3 ft) are on the average less expensive. In the large sizes, plate columns tend to be the more economical.

SUMMARY OF KEY EQUATIONS

The key equations for absorption and stripping calculations for tower height, including a summary of earlier material, are presented below.

For packed tower absorption

$$N_{OG} = \frac{\ln \left[\left(\frac{y_1 - mx_2}{y_2 - mx_2} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{1 - \frac{1}{A}} \quad (10.13)$$

For stripping

$$N_{OL} = \frac{\ln \left[\left\{ \frac{x_2 - (y_1/m)}{x_1 - (y_1/m)} \right\} (1 - A) + A \right]}{1 - A} \quad (10.26)$$

where the subscripts 1 and 2 refer to bottom and top conditions, respectively. In addition, $A = L_m/mG_m$ and $S = 1.0/A$.

To use Figure 10.12 for stripping calculations, replace the y coordinate, x coordinate, and parameter by $[x_1 - (y_1/m)]/[x_2 - (y_2/m)]$, N_{OL} , and S , respectively, where $S = 1.0/A = mG_m/L_m$.

For plate tower absorption

$$N = \frac{\log \left[\left(\frac{y_{N+1} - mx_0}{y_1 - mx_0} \right) \left(1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{\log A} \quad (10.17)$$

Note: The term \ln , rather than \log , may also be employed in both the numerator and denominator.

If A approaches unity, Equation (10.17) becomes

$$N = \frac{y_{N+1} - y_1}{y_1 - mx_0} \quad (10.28)$$

or

$$\frac{y_{N+1} - y_1}{y_{N+1} - mx_0} = \frac{N}{N + 1} \quad (10.23)$$

Note that the subscripts 1 and N refer to the top and bottom of the column, respectively. For stripping in plate towers

$$N = \frac{\log \left[\left\{ \frac{x_0 - (y_{N+1}/m)}{x_N - (y_{N+1}/m)} \right\} \left(1 - \frac{1}{S} \right) + \frac{1}{S} \right]}{\log S} \quad (10.29)$$

or

$$\frac{x_0 - x_N}{x_0 - (y_{N+1}/m)} = \frac{S^{N+1} - S}{S^{N+1} - 1} \quad (10.30)$$

If S is approximately 1.0, one may use either of the following equations:

$$N = \frac{x_0 - x_N}{x_N - (y_{N+1}/m)} \quad (10.31)$$

$$\frac{x_0 - x_N}{x_0 - (y_{N+1}/m)} = \frac{N}{N + 1} \quad (10.32)$$

To use Figure 10.16 for stripping, replace the y coordinate and the parameter A by $[x_N - (y_{N+1}/m)]/[x_0 - (y_{N+1}/m)]$ and S , respectively.

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