

PERRY'S CHEMICAL ENGINEERS' HANDBOOK

佩里化学工程师手册

SEVENTH EDITION

(第七版)

Robert H. Perry
Don W. Green



McGraw-Hill



科学出版社

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Gas Absorption and Gas-Liquid System Design*

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Nomenclature

a_c	Effective interfacial area	m^2/m^3	ft^2/ft^3	L_w	Weir length	m	ft
A	Cross sectional area	m^2	ft^2	m	Slope of equilibrium curve = dy^*/dx	-/-	-/-
A_f	Fractional open area	-/-	-/-	M	Molecular weight	kg/kmol	lb/lb-mol
A	Absorption factor	-/-	-/-	nA	Rate of solute transfer	kmol/s	lb-mol/s
A_e, A'	Effective absorption factor (Edmister)	-/-	-/-	p	Partial pressure	kPa	atm
c	Concentration	kg-moles/ m^3	lb-mol/ ft^3	P, p_T	Total pressure	kPa	atm
c'	Stokes-Cunningham correction factor for terminal settling velocity	-/-	-/-	Q	Volumetric flow rate of liquid	m^3/s	ft^3/s
C_{ab}, C_c	Flooding coefficient	m/s	ft/s	\bar{Q}	Volumetric flow rate of gas or vapor	m^3/s	ft^3/s
C_o	Discharge coefficient	-/-	-/-	R	Gas constant		
d	Diameter	m	ft	R_h	Hydraulic radius	m	ft
d_b	Bubble diameter	m	ft	s	Length of corrugation side, structured packing	m	ft
d_h	Hole diameter	m	ft	S	Stripping factor	-/-	-/-
d_o	Orifice diameter	m	ft	S_e, S'	Effective stripping factor (Edmister)	-/-	-/-
d_{pc}	Cut size of a particle collected in a device, 50% mass efficiency	μm	ft	T	Absolute temperature	K	$^{\circ}F$
d_{pmd}	Mass median size particle in the pollutant gas	μm	ft	TS	Tray or plate spacing	m	ft
$d_{p0.50}$	Aerodynamic diameter of a real median size particle	μm	ft	U	Linear velocity of gas	m/s	ft/s
D	Diffusion coefficient	m^2/s	ft^2/s	U_a	Velocity of gas through active area	m/s	ft/s
D_{32}	Sauter mean diameter	m	ft	U_n	Velocity of gas through net area	m/s	ft/s
D_{vm}	Volume mean diameter	m	ft	U_i	Superficial velocity of gas	m/s	ft/s
e	Entrainment, mass liquid/mass gas	kg/kg	lb/lb	x	Mole fraction, liquid phase	-/-	-/-
E	Plate or stage efficiency, fractional	-/-	-/-	x^*	Liquid mole fraction, equilibrium condition	-/-	-/-
E	Power dissipation per mass	W	btu/lb	y	Mole fraction, gas or vapor phase	-/-	-/-
E_e	Murphree plate efficiency, with entrainment, gas concentrations, fractional	-/-	-/-	y^*	Gas mole fraction, equilibrium condition	-/-	-/-
E_g	Point efficiency, gas phase only, fractional	-/-	-/-	Z	Height, plate spacing	m	ft
E_{oc}	Overall column efficiency, fractional	-/-	-/-	Greek symbols			
E_{op}	Overall point efficiency, gas concentrations, fractional	-/-	-/-	α	Relative volatility	-/-	-/-
E_{mv}	Murphree plate efficiency, gas concentrations, fractional	-/-	-/-	β	Aeration factor	-/-	-/-
f	Fractional approach to flood	-/-	-/-	ϵ	Void fraction	-/-	-/-
F	F-factor for gas loading	$m/s(kg/m^3)^{0.5}$	$ft/s(lb/ft^3)^{0.5}$	ϕ	Relative froth density	-/-	-/-
F_{LG}	Flow parameter	-/-	-/-	ψ	Activity coefficient	-/-	-/-
g	Gravitational constant	m/s^2	ft/s^2	Γ	Flow rate per length	kg/(s-m)	lb/(ft-s)
g_c	Conversion factor	$1.0 (kg \cdot m/N \cdot s^2)$	$32.2 (lb \cdot ft/lb_f \cdot s^2)$	δ	Effective film thickness	m	ft
G	Gas phase mass velocity	kg/s- m^2	lb/hr- ft^2	η	Collection efficiency, fractional	-/-	-/-
G_M	Gas phase molar velocity	kg-moles/s- m^2	lb-mol/hr- ft^2	λ	Stripping factor = $m/(L_{wv}/G_M)$	-/-	-/-
h	Pressure head	mm	ft	μ	Absolute viscosity	Pa-s	lb/(ft-s)
h_f	Height of froth	m	ft	μm	Microns	m	ft
h_T	Height of contacting	m	ft	v	Kinematic viscosity	m^2/s	ft^2/s
H	Henry's law constant			π	3.1416...	-/-	-/-
H'	Henry's law constant			θ	Residence time	s	s
H	Height of a transfer unit	m	ft	ρ	Density	kg/ m^3	lb/ ft^3
H_g	Height of a gas phase transfer unit	m	ft	σ	Surface tension	mN/m	dyn/cm
H_{og}	Height of an overall transfer unit, gas phase concentrations	m	ft	ψ	Fractional entrainment	-/-	-/-
H_{ol}	Height of an overall transfer unit, liquid phase concentrations	m	ft	Subscripts			
H_L	Height of a liquid phase transfer unit	m	ft	A	Species A		
H'	Henry's law coefficient	kPa/mole fraction	atm/mole fraction	AB	Species A diffusing through species B		
HETP	Height equivalent to a theoretical plate or stage	m	ft	B	Species B		
k_1	First order reaction velocity constant	1/s	1/sec	e	Effective value		
k_2	Second order reaction velocity constant	$m^3/(s \cdot kmol)$	$ft^3/(h \cdot lb \cdot mol)$	i	Interface value		
k	Individual phase mass transfer coefficient	m/s	ft/sec	C	Gas or vapor		
k_G	gas phase mass transfer coefficient	m/s	ft/sec	L	Liquid		
k_L	liquid phase mass transfer coefficient	m/s	ft/sec	p	Particle		
K	Vapor-liquid equilibrium ratio	-/-	-/-	w	water		
K_{OG}, K_G	Overall mass transfer coefficient, gas concentrations	m/s	ft/sec	1	Tower bottom		
K_{OL}	Overall mass transfer coefficient, liquid concentrations	m/s	ft/sec	2	Tower top		
L	Liquid mass velocity	kg/ $m^2 \cdot s$	lb/ $ft^2 \cdot s$	Dimensionless Groups			
L_M	Liquid molar mass velocity	kmoles/ $m^2 \cdot s$	lb-mol/ $ft^2 \cdot s$	N_{Fr}	Froude number = $(U_i^2)/(Sg)$		
				N_{Re}	Reynolds number = $(SU_p \rho_C)/(\mu_C)$		
				N_{Sc}	Schmidt number = $\mu/(\rho D)$		
				N_{We}	Weber number = $(U_i^2 \rho_L S)/(\sigma_g)$		

GENERAL REFERENCES: Astarita, G., *Mass Transfer with Chemical Reaction*, Elsevier, New York, 1967. Astarita, G., D. W. Savage and A. Bisio, *Gas Treating with Chemical Solvents*, Wiley, New York, 1983. Billet, R., *Distillation Engineering*, Chemical Publishing Co., New York, 1979. Danckwerts, P. V., *Gas-Liquid Reactions*, McGraw-Hill, New York, 1970. *Distillation and Absorption 1987*, Rugby, U.K., Institution of Chemical Engineers, 1988. *Distillation and Absorption 1992*, Rugby, U.K., Institution of Chemical Engineers, 1992. Hines,

A. L. and R. N. Maddox, *Mass Transfer—Fundamentals and Applications*, Prentice Hall, Englewood Cliffs, New Jersey, 1985. Kister, H. Z., *Distillation Design*, McGraw-Hill, New York, 1992. Lockett, M. J., *Distillation Tray Fundamentals*, Cambridge, U.K., Cambridge University Press, 1986. Kohl, A. L. and F. C. Riesenfeld, *Gas Purification*, 4th ed., Gulf, Houston, 1985. Sherwood, T. K., R. L. Pigford, C. R. Wilke, *Mass Transfer*, McGraw-Hill, New York, 1975. Treybal, R. E., *Mass Transfer Operations*, McGraw-Hill, New York, 1980.

INTRODUCTION

Definitions Gas absorption is a unit operation in which soluble components of a gas mixture are dissolved in a liquid. The inverse operation, called stripping or desorption, is employed when it is desired to transfer volatile components from a liquid mixture into a gas. Both absorption and stripping, in common with distillation (Sec. 13), make use of special equipment for bringing gas and liquid phases into intimate contact. This section is concerned with the design of gas-liquid contacting equipment, as well as with the design of absorption and stripping processes.

Equipment Absorption, stripping, and distillation operations are usually carried out in vertical, cylindrical columns or towers in which devices such as plates or packing elements are placed. The gas and liquid normally flow countercurrently, and the devices serve to provide the contacting and development of interfacial surface through which mass transfer takes place. Background material on this mass transfer process is given in Sec. 5.

Design Procedures The procedures to be followed in specifying the principal dimensions of gas absorption and distillation equipment are described in this section and are supported by several worked-out examples. The experimental data required for executing the designs are keyed to appropriate references or to other sections of the handbook.

For absorption, stripping, and distillation, there are three main steps involved in design:

1. *Data on the gas-liquid or vapor-liquid equilibrium for the system at hand.* If absorption, stripping, and distillation operations are considered equilibrium-limited processes, which is the usual approach, these data are critical for determining the maximum possible separation. In some cases, the operations are considered rate-based (see Sec. 13) but require knowledge of equilibrium at the phase interface. Other data required include physical properties such as viscosity and density and thermodynamic properties such as enthalpy. Section 2 deals with sources of such data.

2. *Information on the liquid- and gas-handling capacity of the contacting device chosen for the particular separation problem.* Such information includes pressure drop characteristics of the device, in order that an optimum balance between capital cost (column cross section) and energy requirements might be achieved. Capacity and pressure drop characteristics of the available devices are covered later in this Sec. 14.

3. *Determination of the required height of contacting zone for the separation to be made as a function of properties of the fluid mixtures and mass-transfer efficiency of the contacting device.* This determination involves the calculation of mass-transfer parameters such as heights of transfer units and plate efficiencies as well as equilibrium or rate parameters such as theoretical stages or numbers of transfer units. An additional consideration for systems in which chemical reaction occurs is the provision of adequate residence time for desired reactions to occur, or minimal residence time to prevent undesired reactions from occurring. For equilibrium-based operations, the parameters for required height are covered in the present section.

Data Sources in the Handbook Sources of data for the analysis or design of absorbers, strippers, and distillation columns are mani-

fold, and a detailed listing of them is outside the scope of the presentation in this section. Some key sources within the handbook are shown in Table 14-1.

Equilibrium Data Finding reliable gas-liquid and vapor-liquid equilibrium data may be the most time-consuming task associated with the design of absorbers and other gas-liquid contactors, and yet it may be the most important task at hand. For gas solubility, an important data source is the set of volumes edited by Kertes et al., *Solubility Data Series*, published by Pergamon Press (1979 ff.). In the introduction to each volume, there is an excellent discussion and definition of the various methods by which gas solubility data have been reported, such as the Bunsen coefficient, the Kuenen coefficient, the Ostwald coefficient, the absorption coefficient, and the Henry's law coefficient. The fourth edition of *The Properties of Gases and Liquids* by Reid, Prausnitz and Poling (McGraw-Hill, New York, 1987) provides data and recommended estimation methods for gas solubility as well as the broader area of vapor-liquid equilibrium. Finally, the Chemistry Data Series by Gmehling et al., especially the title *Vapor-Liquid Equilibrium Collection* (DECHEMA, Frankfurt, Germany, 1979 ff.), is a rich source of data evaluated against the various models used for interpolation and extrapolation. Section 13 of this handbook presents a good discussion of equilibrium *K* values.

TABLE 14-1 Directory to Key Data for Absorption and Gas-Liquid Contactor Design

Type of data	Section
Phase equilibrium data	
Gas solubilities	2
Pure component vapor pressures	2
Equilibrium <i>K</i> values	13
Thermal data	
Heats of solution	2
Specific heats	2
Latent heats of vaporization	2
Transport property data	
Diffusion coefficients	
Liquids	2
Gases	2
Viscosities	
Liquids	2
Gases	2
Densities	
Liquids	2
Gases	2
Surface tensions	2
Packed tower data	
Pressure drop and flooding	14
Mass transfer coefficients	5
HTU, physical absorption	5
HTU with chemical reaction	14
Height equivalent to a theoretical plate (HETP)	
Plate tower data	
Pressure drop and flooding	14
Plate efficiencies	14
Costs of gas-liquid contacting equipment	14

DESIGN OF GAS-ABSORPTION SYSTEMS

General Design Procedure The designer ordinarily is required to determine (1) the best solvent; (2) the best gas velocity through the absorber, namely the vessel diameter; (3) the height of the vessel and its internal members, which is the height and type of packing or the number of contacting trays; (4) the optimum solvent circulation through the absorber and stripper; (5) the temperatures of streams entering and leaving the absorber and the quantity of heat to be removed to account for heat of solution and other thermal effects; (6) the pressures at which the absorber and stripper will operate; and (7) the mechanical design of the absorption and stripping vessels (normally columns or towers), including flow distributors, packing supports, and so on. This section is concerned with all these choices.

The problem presented to the designer of a gas-absorption unit usually specifies the following quantities: (1) gas flow rate; (2) gas composition, at least with respect to the component or components to be absorbed; (3) operating pressure and allowable pressure drop across the absorber; (4) minimum degree of recovery of one or more solutes; and, possibly, (5) the solvent to be employed. Items 3, 4, and 5 may be subject to economic considerations and therefore are sometimes left up to the designer. For determining the number of variables that must be specified in order to fix a unique solution for the design of an absorber one can use the same phase-rule approach described in Sec. 13 for distillation systems.

Recovery of the solvent, sometimes by chemical means but more often by distillation, is almost always required, and the recovery system ordinarily is considered an integral part of the absorption-system process design. A more efficient solvent-stripping operation normally will result in a less costly absorber because of a smaller concentration of residual dissolved solute in the regenerated solvent; however, this may increase the overall cost of solvent recovery. A more detailed discussion of these and other economic considerations is presented later in this section.

Selection of Solvent When choice is possible, preference is given to liquids with high solubilities for the solute; a high solubility reduces the amount of solvent to be circulated. The solvent should be relatively nonvolatile, inexpensive, noncorrosive, stable, nonviscous, nonfoaming, and preferably nonflammable. Since the exit gas normally leaves saturated with solvent, solvent loss can be costly and may present environmental contamination problems. Thus, low-cost solvents may be chosen over more expensive ones of higher solubility or lower volatility.

Water generally is used for gases fairly soluble in water, oils for light hydrocarbons, and special chemical solvents for acid gases such as CO_2 , SO_2 , and H_2S . Sometimes a reversible chemical reaction will result in a very high solubility and a minimum solvent rate. Data on actual systems are desirable when chemical reactions are involved, and those available are referenced later under "Absorption with Chemical Reaction."

Selection of Solubility Data Solubility values determine the liquid rate necessary for complete or economic solute recovery and so are essential to design. Equilibrium data generally will be found in one of three forms: (1) solubility data expressed either as solubility in weight or mole percent or as Henry's-law coefficients, (2) pure-component vapor pressures, or (3) equilibrium distribution coefficients (K values). Data for specific systems may be found in Sec. 2; additional references to sources of data are presented in this section.

In order to define completely the solubility of a gas in a liquid, it generally is necessary to state the temperature, the equilibrium partial pressure of the solute gas in the gas phase, and the concentration of the solute gas in the liquid phase. Strictly speaking, the total pressure on the system also should be stated, but for low total pressures, less than about 507 kPa (5 atm), the solubility for a particular partial pressure of solute gas normally will be relatively independent of the total pressure of the system.

For dilute concentrations of many gases and over a fairly wide range for some gases, the equilibrium relationship is given by Henry's law, which relates the partial pressure developed by a dissolved solute A in a liquid solvent B by one of the following equations:

$$p_A = Hx_A \quad (14-1)$$

$$p_A = H'c_A \quad (14-2)$$

where H is the Henry's law coefficient expressed in kilopascals per mole-fraction solute in liquid and H' is the Henry's law coefficient expressed in kilopascals per kilomole per cubic meter.

Although quite useful when it can be applied, this law should be checked experimentally to determine the accuracy with which it can be used. If Henry's law holds, the solubility is defined by stating the value of the constant H (or H') along with the temperature and the solute partial pressure for which it is to be employed.

For quite a number of gases, Henry's law holds very well when the partial pressure of the solute is less than about 100 kPa (1 atm). For partial pressures of the solute gas greater than 100 kPa, H seldom is independent of the partial pressure of the solute gas, and a given value of H can be used over only a narrow range of partial pressures. There is a strongly nonlinear variation of Henry's-law constants with temperature as discussed by Schulze and Prausnitz [*Ind. Eng. Chem. Fundam.*, **20**, 175 (1981)]. Consultation of this reference is recommended before considering temperature extrapolations of Henry's-law data.

Additional data and information on the applicability of Henry's-law constants can be found in the references cited earlier in the subsection "Directory to Key Gas-Absorption Data." The use of Henry's-law constants is illustrated by the following examples.

Example 1: Gas Solubility It is desired to find out how much hydrogen can be dissolved in 100 weights of water from a gas mixture when the total pressure is 101.3 kPa (760 torr; 1 atm), the partial pressure of the H_2 is 26.7 kPa (200 torr), and the temperature is 20°C. For partial pressures up to about 100 kPa the value of H is given in Sec. 3 as 6.92×10^6 kPa (6.83×10^8 atm) at 20°C. According to Henry's law,

$$x_{\text{H}_2} = p_{\text{H}_2}/H_{\text{H}_2} = 26.7/6.92 \times 10^6 = 3.86 \times 10^{-6}$$

The mole fraction x is the ratio of the number of moles of H_2 in solution to the total moles of all constituents contained. To calculate the weights of H_2 per 100 weights of H_2O , one can use the following formula, where the subscripts A and w correspond to the solute (hydrogen) and solvent (water):

$$\begin{aligned} \left(\frac{x_A}{1-x_A} \right) \frac{M_A}{M_w} 100 &= \left(\frac{3.86 \times 10^{-6}}{1-3.86 \times 10^{-6}} \right) \frac{2.02}{18.02} 100 \\ &= 4.33 \times 10^{-5} \text{ weights } \text{H}_2/100 \text{ weights } \text{H}_2\text{O} \\ &= 0.43 \text{ parts per million weight} \end{aligned}$$

Pure-component vapor pressures can be used for predicting solubilities for systems in which **Raoult's law** is valid. For such systems $p_A = p_A^0 x_A$, where p_A^0 is the pure-component vapor pressure of the solute and p_A is its partial pressure. Extreme care should be exercised when attempting to use pure-component vapor pressures to predict gas-absorption behavior. Both liquid-phase and vapor-phase nonidealities can cause significant deviations from the behavior predicted from pure-component vapor pressures in combination with Raoult's law. Vapor-pressure data are available in Sec. 3 for a variety of materials.

Whenever data are available for a given system under similar conditions of temperature, pressure, and composition, **equilibrium distribution coefficients** ($K = y/x$) provide a much more reliable tool for predicting vapor-liquid distributions. A detailed discussion of equilibrium K values is presented in Sec. 13.

Calculation of Liquid-to-Gas Ratio The minimum possible liquid rate is readily calculated from the composition of the entering gas and the solubility of the solute in the exit liquor, saturation being assumed. It may be necessary to estimate the temperature of the exit liquid based on the heat of solution of the solute gas. Values of latent and specific heats and values of heats of solution (at infinite dilution) are given in Sec. 2.

The actual liquid-to-gas ratio (solvent-circulation rate) normally will be greater than the minimum by as much as 25 to 100 percent and may be arrived at by economic considerations as well as by judgment and experience. For example, in some packed-tower applications involving very soluble gases or vacuum operation, the minimum quantity of solvent needed to dissolve the solute may be insufficient to keep the packing surface thoroughly wet, leading to poor distribution of the liquid stream.

When the solute concentration in the inlet gas is low and when nearly all the solute is being absorbed (this is the usual case), the approximation

$$y_1 G_M \approx x_1 L_M \approx (y_1^*/m) L_M \quad (14-3)$$

leads to the conclusion that the ratio mG_M/L_M represents the fractional approach of the exit liquid to saturation with the inlet gas, i.e.,

$$mG_M/L_M \approx y_1^*/y_1 \quad (14-4)$$

Optimization of the liquid-to-gas ratio in terms of total annual costs often suggests that the molar liquid-to-gas ratio L_M/G_M should be about 1.2 to 1.5 times the theoretical minimum corresponding to equilibrium at the rich end of the tower (infinite height), provided flooding is not a problem. This would be an alternative to assuming that $L_M/G_M \approx m/0.7$, for example.

When the exit-liquor temperature rises owing to the heat of absorption of the solute, the value of m changes through the tower, and the liquid-to-gas ratio must be chosen to give reasonable values of $m_1 G_M/L_M$ and $m_2 G_M/L_M$, where the subscripts 1 and 2 refer to the bottom and top of the absorption tower respectively. For this case the value of $m_2 G_M/L_M$ will be taken to be somewhat less than 0.7, so that the value of $m_1 G_M/L_M$ will not approach unity too closely. This rule-of-thumb approach is useful only when low solute concentrations and mild heat effects are involved.

When the solute has a large heat of solution or when the feed gas contains high percentages of the solute, one should consider the use of internal cooling coils or intermediate external heat exchangers in a plate-type tower to remove the heat of absorption. In a packed tower, one could consider the use of multiple packed sections with intermediate liquid-withdrawal points so that the liquid could be cooled by external heat exchange.

Selection of Equipment Packed columns usually are chosen for very corrosive materials, for liquids that foam badly, for either small- or large-diameter towers involving very low allowable pressure drops, and for small-scale operations requiring diameters of less than 0.6 m (2 ft). The type of packing is selected on the basis of resistance to corrosion, mechanical strength, capacity for handling the required flows, mass-transfer efficiency, and cost. Economic factors are discussed later in this section.

Plate columns may be economically preferable for large-scale operations and are needed when liquid rates are so low that packing would be inadequately wetted, when the gas velocity is so low (owing to a very high L/G) that axial dispersion or "pumping" of the gas back down the (packed) column can occur, or when intermediate cooling is desired. Also, plate towers may have a better turndown ratio and are less subject to fouling by solids than are packed towers. Details on the operating characteristics of plate towers are given later in this section.

Column Diameter and Pressure Drop Flooding determines the minimum possible diameter of the absorber column, and the usual design is for 60 to 80 percent of the flooding velocity. Maximum allowable pressure drop may be determined by the cost of energy for compression of the feed gas. For systems having a significant tendency to foam, the maximum allowable velocity will be lower than estimated flooding velocity, especially for plate towers. The safe range of operating velocities should include the velocity one would derive from economic considerations, as discussed later. Methods for predicting flooding velocities and pressure drops are given later in this section.

Computation of Tower Height The required height of a gas-absorption or stripping tower depends on (1) the phase equilibria involved, (2) the specified degree of removal of the solute from the gas, and (3) the mass-transfer efficiency of the apparatus. These same considerations apply both to plate towers and to packed towers. Items 1 and 2 dictate the required number of theoretical stages (plate tower) or transfer units (packed tower). Item 3 is derived from the tray efficiency and spacing (plate tower) or from the height of one transfer unit (packed tower). Solute-removal specifications normally are derived from economic considerations.

For plate towers, the approximate design methods described below may be used in estimating the number of theoretical stages, and the tray efficiencies and spacings for the tower can be specified on the basis of the information given later. Considerations involved in the rigorous design of theoretical stages for plate towers are treated in Sec. 13.

For packed towers, the continuous differential nature of the contact between gas and liquid leads to a design procedure involving the solution of differential equations, as described in the next subsection.

It should be noted that the design procedures discussed in this section are not applicable to reboiled absorbers, which should be designed according to the methods described in Sec. 13.

Caution is advised in distinguishing between systems involving pure physical absorption and those in which a chemical reaction can significantly affect design procedures.

Selection of Stripper-Operating Conditions Stripping involves the removal of one or more volatile components from a liquid by contacting it with a gas such as steam, nitrogen, or air. The operating conditions chosen for stripping normally result in a low solubility of the solute (i.e., a high value of m), so that the ratio mG_M/L_M will be larger than unity. A value of 1.4 may be used for rule-of-thumb calculations involving pure physical desorption. For plate-tower calculations the stripping factor $S = KG_M/L_M$, where $K = y^*/x$, usually is specified for each tray.

When the solvent from an absorption operation must be regenerated for recycling back to the absorber, one may employ a "pressure-swing concept," a "temperature-swing concept," or a combination of both in specifying stripping conditions. In pressure-swing operation the temperature of the stripper is about the same as that of the absorber, but the stripping pressure is much lower. In temperature-swing operation the pressures are about equal, but the stripping temperature is much higher than the absorption temperature.

In pressure-swing operation a portion of the dissolved gas may be "sprung" from the liquid by the use of a flash drum upstream of the stripping-tower feed point. This type of operation is discussed by Burrows and Preece [*Trans. Inst. Chem. Eng.*, **32**, 99 (1954)] and by Langley and Haselden [*Inst. Chem. Eng. Symp. Ser. (London)*, no. 28 (1968)]. If the flashing of the feed liquid takes place inside the stripping tower, this effect must be accounted for in the design of the upper section in order to avoid overloading and flooding near the top of the tower.

More often than not the rate at which residual absorbed gas can be driven from the liquid in a stripping tower is limited by the rate of a chemical reaction, in which case the liquid-phase residence time (and hence, the tower liquid holdup) becomes the most important design factor. Thus, many stripper-regenerators are designed on the basis of liquid holdup rather than on the basis of mass transfer rate.

Approximate design equations applicable only to the case of pure physical desorption are developed later in this section for both packed and plate stripping towers. A more rigorous approach using distillation concepts may be found in Sec. 13. A brief discussion of desorption with chemical reaction is given in the subsection "Absorption with Chemical Reaction."

Design of Absorber-Stripper Systems The solute-rich liquor leaving a gas absorber normally is distilled or stripped to regenerate the solvent for recirculation back to the absorber, as depicted in Fig. 14-1. It is apparent that the conditions selected for the absorption step (e.g., temperature, pressure, L_M/G_M) will affect the design of the stripping tower, and, conversely, a selection of stripping conditions will affect the absorber design. The choice of optimum operating conditions for an absorber-stripper system therefore involves a combination of economic factors and practical judgments as to the operability of the system within the context of the overall process flow sheet. Note that in Fig. 14-1 the stripping vapor is provided by a reboiler; alternatively, an extraneous stripping gas may be used.

An appropriate procedure for executing the design of an absorber-stripper system is to set up a carefully selected series of design cases and then evaluate the investment costs, the operating costs, and the operability of each case. Some of the economic factors that need to be considered in selecting the optimum absorber-stripper design are discussed later in the subsection "Economic Design of Absorption Systems."

Importance of Design Diagrams One of the first things a designer should try to do is lay out a carefully constructed equilibrium curve, $y^* = F(x)$, on an xy diagram, as shown in Fig. 14-2. A horizontal line corresponding to the inlet-gas composition y_1 is then the locus of feasible outlet-liquor compositions, and a vertical line corresponding to the inlet-solvent-liquor composition x_2 is the locus of feasible out-

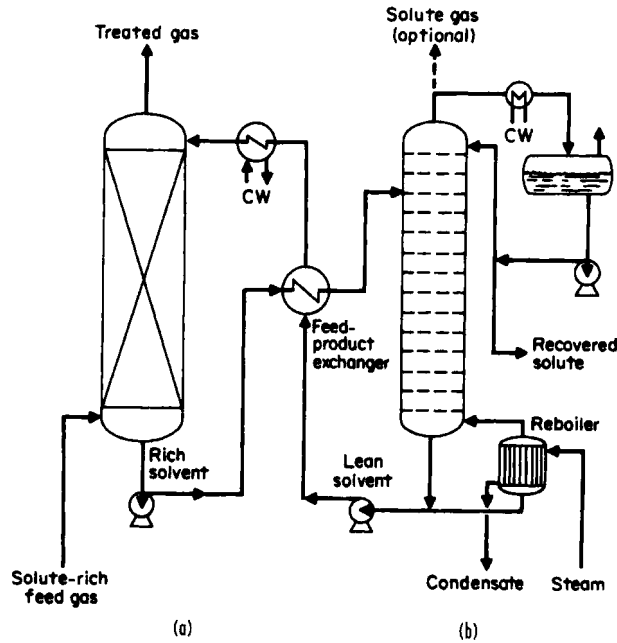


FIG. 14-1 Gas absorber using a solvent regenerated by stripping. (a) Absorber. (b) Stripper.

let-gas compositions. These lines are indicated as $y = y_1$ and $x = x_2$ respectively on Fig. 14-2.

For gas absorption, the region of feasible operating lines lies above the equilibrium curve; for stripping, the feasible region for operating lines lies below the equilibrium curve. These feasible regions are bounded by the equilibrium curve and by the lines $x = x_2$ and $y = y_1$. By inspection, one should be able to visualize those operating lines that are feasible and those that would lead to "pinch points" within the tower. Also, it is possible to determine if a particular proposed design for solute recovery falls within the feasible envelope.

Once the design recovery for an absorber has been established, the operating curve can be constructed by first locating the point x_2, y_2 on the diagram. The intersection of the horizontal line corresponding to the inlet gas composition y_1 with the equilibrium curve $y^o = F(x)$

defines the theoretical minimum liquid-to-gas ratio for systems in which there are no intermediate pinch points. The operating line which connects this point with the point x_2, y_2 corresponds to the minimum value of L_M/G_M . The actual design value of L_M/G_M normally should be around 1.2 to 1.5 times this minimum. Thus, the actual design operating line for a gas absorber will pass through the point x_2, y_2 and will intersect the line $y = y_1$ to the left of the equilibrium curve.

For stripping one begins by using the design specification to locate the point x_1, y_1 . Then the intersection of the vertical line $x = x_2$ with the equilibrium curve $y^o = F(x)$ defines the theoretical minimum gas-to-liquid ratio. The actual value of G_M/L_M is chosen to be about 20 to 50 percent higher than this minimum, so the actual design operating line will intersect the line $x = x_2$ at a point somewhat below the equilibrium curve.

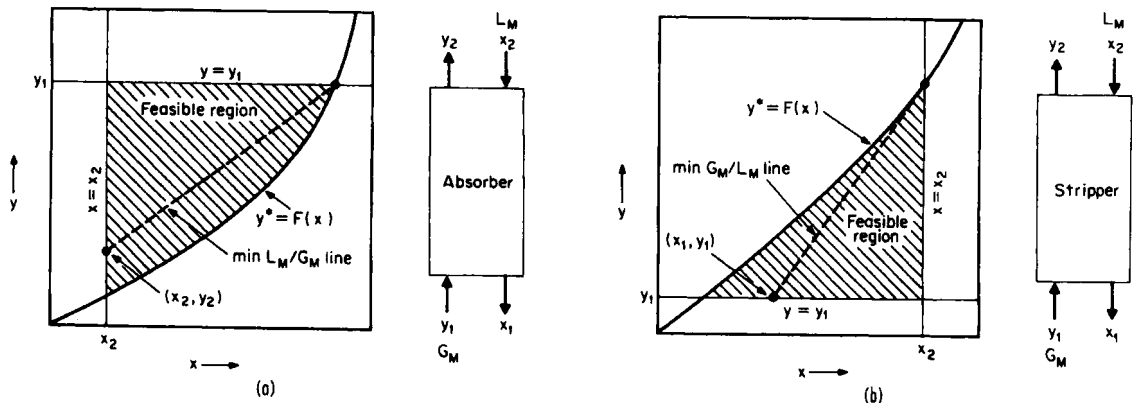


FIG. 14-2 Design diagrams for (a) absorption and (b) stripping.

14-8 GAS ABSORPTION AND GAS-LIQUID SYSTEM DESIGN

Design diagrams minimize the possibility of making careless mistakes and allow one to assess easily the effects of operating variable changes on the operability of the system relative to pinch points, etc. Whenever analytical calculations or computer programs are being used for the design of gas-absorption systems, the construction of

design diagrams based either on calculation results or on computer printouts may reveal problem areas or even errors in the design concept. It is strongly recommended that design diagrams be employed whenever possible.

PACKED-TOWER DESIGN

Methods for estimating the height of the active section of **counter-flow differential contactors** such as packed towers, spray towers, and falling-film absorbers are based on rate expressions representing mass transfer at a point on the gas-liquid interface and on material balances representing the changes in bulk composition in the two phases that flow past each other. The rate expressions are based on the interphase mass-transfer principles described in Sec. 5. Combination of such expressions leads to an integral expression for the number of transfer units or to equations related closely to the number of theoretical plates. The paragraphs which follow set forth convenient methods for using such equations, first in a general case and then for cases in which simplifying assumptions are valid.

Use of Mass-Transfer-Rate Expression Figure 14-3 shows a section of a packed absorption tower together with the nomenclature that will be used in developing the equations which follow. In a differential section dh , we can equate the rate at which solute is lost from the gas phase to the rate at which it is transferred through the gas phase to the interface as follows:

$$-d(G_M y) = -G_M dy - y dG_M = N_A a dh \quad (14-5)$$

When only one component is transferred,

$$dG_M = -N_A a dh \quad (14-6)$$

Substitution of this relation into Eq. (14-5) and rearranging yields

$$dh = -\frac{G_M dy}{N_A a(1-y)} \quad (14-7)$$

For this derivation we use the gas-phase rate expression $N_A = k_G(y - y_i)$ and integrate over the tower to obtain

$$h_T = \int_{y_2}^{y_1} \frac{G_M dy}{k_G a(1-y)(y-y_i)} \quad (14-8)$$

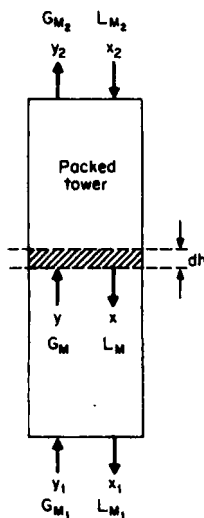


FIG. 14-3 Nomenclature for material balances in a packed-tower absorber or stripper.

Multiplying and dividing by y_{BM} place Eq. (14-8) into the $H_G N_G$ format

$$h_T = \int_{y_2}^{y_1} \left[\frac{G_M}{k_G a y_{BM}} \right] \frac{y_{BM} dy}{(1-y)(y-y_i)} \\ = H_{G,w} \int_{y_2}^{y_1} \frac{y_{BM} dy}{(1-y)(y-y_i)} = H_{G,w} N_G \quad (14-9)$$

The general expression given by Eq. (14-8) is more complex than normally is required, but it must be used when the mass-transfer coefficient varies from point to point, as may be the case when the gas is not dilute or when the gas velocity varies as the gas dissolves. The values of y_i to be used in Eq. (14-8) depend on the local liquid composition x_i and on the temperature. This dependency is best represented by using the operating and equilibrium lines as discussed later.

Example 3 illustrates the use of Eq. (14-8) for scrubbing chlorine from air with aqueous caustic solution. For this case one can make the simplifying assumption that y_i , the interfacial partial pressure of chlorine over the aqueous caustic solution, is zero owing to the rapid and complete reaction of the chlorine after it dissolves. We note that the feed gas is not dilute.

Example 2: Packed Height Requirement Let us compute the height of packing needed to reduce the chlorine concentration of 0.537 kg/(s·m²), or 396 lb/(h·ft²), of a chlorine-air mixture containing 0.503 mole-fraction chlorine to 0.0403 mole fraction. On the basis of test data described by Sherwood and Pigford (*Absorption and Extraction*, McGraw-Hill, 1952, p. 121) the value of $k_G a y_{BM}$ at a gas velocity equal to that at the bottom of the packing is equal to 0.1175 kmol/(s·m²), or 26.4 lb-mol/(h·ft²). The equilibrium back pressure y_i can be assumed to be negligible.

Solution. By assuming that the mass-transfer coefficient varies as the 0.8 power of the local gas mass velocity, we can derive the following relation:

$$\tilde{k}_G a = k_G a y_{BM} = 0.1175 \left[\frac{71y + 29(1-y)}{71y_1 + 29(1-y_1)} \left(\frac{1-y_1}{1-y} \right) \right]^{0.8}$$

where 71 and 29 are the molecular weights of chlorine and air respectively. Noting that the inert-gas (air) flow rate is given by $G_M' = G_M(1-y) = 5.34 \times 10^{-3}$ kmol/(s·m²), or 3.94 lb-mol/(h·ft²), and introducing these expressions into the integral gives

$$h_T = 1.82 \int_{0.0403}^{0.503} \left[\frac{1-y}{29+42y} \right]^{0.8} \frac{dy}{(1-y)^2 \ln[1/(1-y)]}$$

This definite integral can be evaluated numerically by the use of Simpson's rule to obtain $h_T = 0.305$ m (1 ft).

Use of Operating Curve Frequently, it is not possible to assume that $y_i = 0$ as in Example 2, owing to diffusional resistance in the liquid phase or to the accumulation of solute in the liquid stream. When the back pressure cannot be neglected, it is necessary to supplement the equations with a material balance representing the operating line or curve. In view of the countercurrent flows into and from the differential section of packing shown in Fig. 14-3, a steady-state material balance leads to the following equivalent relations:

$$d(G_M y) = d(L_M x) \quad (14-10)$$

$$G_M' \frac{dy}{(1-y)^2} = L_M' \frac{dx}{(1-x)^2} \quad (14-11)$$

where L_M' = molar mass velocity of the inert-liquid component and G_M' = molar mass velocity of the inert gas. L_M , L_M' , G_M , and G_M' are superficial velocities based on the total tower cross section.

Equation (14-11) is the differential equation of the operating curve, and its integral around the upper portion of the packing is the equation for the operating curve

$$G'_M \left[\frac{y}{1-y} - \frac{y_2}{1-y_2} \right] = L'_M \left[\frac{x}{1-x} - \frac{x_2}{1-x_2} \right] \quad (14-12)$$

For dilute solutions in which the mole fractions of x and y are small, the total molar flows G_M and L_M will be very nearly constant, and the operating-curve equation is

$$G_M(y - y_2) = L_M(x - x_2) \quad (14-13)$$

This equation gives the relation between the bulk compositions of the gas and liquid streams at each level in the tower for conditions in which the operating curve can be approximated by a straight line.

Figure 14-4 shows the relationship between the operating curve and the equilibrium curve $y_i = F(x_i)$ for a typical example involving solvent recovery, where y_i and x_i are the interfacial compositions (assumed to be in equilibrium). Once y is known as a function of x along the operating curve, y_i can be found at corresponding points on the equilibrium curve by

$$(y - y_i)/(x_i - x) = k_L/k_G = k'_L \bar{p}_L/k'_G p_T = L_M H_C / G_M H_L \quad (14-14)$$

where L_M = molar liquid mass velocity, G_M = molar gas mass velocity, H_L = height of one transfer unit based on liquid-phase resistance, and H_C = height of one transfer unit based on gas-phase resistance. Thence, the integral in Eq. (14-8) can be evaluated.

Calculation of Transfer Units In the general case the equations described above must be employed in calculating the height of packing required for a given separation. However, if the local mass-transfer coefficient $k_G a y_{BM}$ is approximately proportional to the first power of the local gas velocity G_M , then the height of one gas-phase transfer unit, defined as $H_C = G_M / k_G a y_{BM}$, will be constant in Eq. (14-9). Similar considerations lead to an assumption that the height of one overall gas-phase transfer unit H_{OC} may be taken as constant. The height of packing required is then calculated according to the relation

$$h_T = H_C N_C = H_{OC} N_{OC} \quad (14-15)$$

where N_C = number of gas-phase transfer units and N_{OC} = number of overall gas-phase transfer units. When H_C and H_{OC} are not constant, it may be valid to employ averaged values between the top and bottom of the tower and the relation

$$h_T = H_{C,av} N_C = H_{OC,av} N_{OC} \quad (14-16)$$

In these equations, the terms N_C and N_{OC} are defined by

$$N_C = \int_{y_2}^{y_1} \frac{y_{BM} dy}{(1-y)(y-y_i)} \quad (14-17)$$

and by

$$N_{OC} = \int_{y_2}^{y_1} \frac{y_{BM}^o dy}{(1-y)(y-y^o)} \quad (14-18)$$

respectively.

Equation (14-18) is the more useful one in practice: it requires either actual experimental H_{OC} data or values estimated by combining

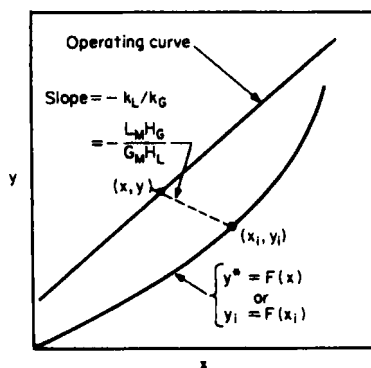


FIG. 14-4 Relationship between equilibrium curve and operating curve in a packed absorber; computation of interfacial compositions.

individual measurements of H_C and H_L by Eq. (14-19). Correlations for predicting H_C , H_L , and H_{OC} in nonreacting systems are presented in Sec. 5.

$$H_{OC} = \frac{y_{BM}}{y_{BM}^o} H_C + \frac{m G_M}{L_M} \frac{x_{BM}}{y_{BM}^o} H_L \quad (14-19a)$$

$$H_{OL} = \frac{x_{BM}}{x_{BM}^o} H_L + \frac{L_M}{m G_M} \frac{y_{BM}}{x_{BM}^o} H_C \quad (14-19b)$$

On occasion the changes in gas flow and in the mole fraction of inert gas are so small that the inclusion of terms such as $(1-y)$ and y_{BM}^o can be neglected or at least can be included in an approximate way. This leads to some of the simplified procedures described later.

One such simplification was suggested by Wiegand [*Trans. Am. Inst. Chem. Eng.*, **36**, 679 (1940)], who pointed out that the logarithmic-mean mole fraction of inert gas y_{BM}^o (or y_{BM}) is often very nearly equal to the arithmetic mean. Thus, substitution of the relation

$$\frac{y_{BM}^o}{(1-y)} = \frac{(1-y^o) + (1-y)}{2(1-y)} = \frac{y-y^o}{2(1-y)} + 1 \quad (14-20)$$

into the equations presented earlier leads to the simplified forms

$$N_C = \frac{1}{2} \ln \frac{1-y_2}{1-y_1} + \int_{y_2}^{y_1} \frac{dy}{y-y_i} \quad (14-21)$$

$$N_{OC} = \frac{1}{2} \ln \frac{1-y_2}{1-y_1} + \int_{y_2}^{y_1} \frac{dy}{y-y^o} \quad (14-22)$$

The second (integral) terms represent the numbers of transfer units for an infinitely dilute gas. The first terms, frequently amounting to only small corrections, give the effect of a finite level of gas concentration.

The procedure for applying Eqs. (14-21) and (14-22) involves two steps: (1) evaluation of the integrals and (2) addition of the correction corresponding to the first (logarithmic) term. The discussion which follows deals only with the evaluation of the integral terms (first step).

The simplest possible case occurs when (1) both the operating and the equilibrium lines are straight (i.e., there are dilute solutions), (2) Henry's law is valid ($y^o/x = y_i/x_i = m$), and (3) absorption heat effects are negligible. Under these conditions, the integral term in Eq. (14-20) may be computed by Colburn's equation [*Trans. Am. Inst. Chem. Eng.*, **35**, 211 (1939)]:

$$N_{OC} = \frac{1}{1 - (m G_M / L_M)} \ln \left[\left(1 - \frac{m G_M}{L_M} \right) \left(\frac{y_1 - m x_2}{y_2 - m x_2} \right) + \frac{m G_M}{L_M} \right] \quad (14-23)$$

Figure (14-5) is a plot of Eq. (14-23) from which the value of N_{OC} can be read directly as a function of $m G_M / L_M$ and the ratio of concentrations. This plot and Eq. (14-23) are equivalent to the use of a logarithmic mean of terminal driving forces, but they are more convenient because one does not need to compute the exit-liquor concentration x_1 .

In many practical situations involving nearly complete cleanup of the gas, an approximate result can be obtained from the equations just presented even when solutions are concentrated or when absorption heat effects are present. In such cases the driving forces in the upper part of the tower are very much smaller than those at the bottom, and the value of $m G_M / L_M$ used in the equations should be the ratio of the slopes of the equilibrium line m and the operating line L_M / G_M in the low-concentration range near the top of the tower.

Another approach is to divide the tower arbitrarily into a lean section (near the top), where approximate methods are valid, and to deal with the rich section separately. If the heat effects in the rich section are appreciable, consideration could be given to installing cooling units near the bottom of the tower. In any event a design diagram showing the operating and equilibrium curves should be prepared to check on the applicability of any simplified procedure. Figure 14-8, presented in Example 6 is one such diagram for an adiabatic absorption tower.

Stripping Equations Stripping, or desorption, involves the removal of a volatile component from the liquid stream by contact with an inert gas such as nitrogen or steam. In this case the change in concentration of the liquid stream is of prime importance, and it

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is more convenient to formulate the rate equation analogous to Eq. (14-6) in terms of the liquid composition x . This leads to the following equations defining numbers of transfer units and heights of transfer units based on liquid-phase resistance:

$$h_T = H_L \int_{x_2}^{x_1} \frac{x_{BM} dx}{(1-x)(x_1-x)} = H_L N_L \quad (14-24)$$

$$h_T = H_{OL} \int_{x_2}^{x_1} \frac{x_{BM}^o dx}{(1-x)(x^o-x)} = H_{OL} N_{OL} \quad (14-25)$$

where, as before, subscripts 1 and 2 refer to the bottom and top of the tower respectively (see Fig. 14-3).

In situations in which one cannot assume that H_L and H_{OL} are constant, these terms must be incorporated inside the integrals in Eqs. (14-24) and (14-25), and the integrals must be evaluated graphically or numerically (by using Simpson's rule, for example). In the normal case involving stripping without chemical reactions, the liquid-phase resistance will dominate, making it preferable to use Eq. (14-25) in conjunction with the relation $H_L \doteq H_{OL}$.

The Wiegand approximations of the above integrals in which arithmetic means are substituted for the logarithmic means x_{BM} and x_{BM}^o are

$$N_L = \frac{1}{2} \ln \frac{1-x_1}{1-x_2} + \int_{x_1}^{x_2} \frac{dx}{x-x_1} \quad (14-26)$$

$$N_{OL} = \frac{1}{2} \ln \frac{1-x_1}{1-x_2} + \int_{x_1}^{x_2} \frac{dx}{x-x^o} \quad (14-27)$$

In these equations, the first term is a correction for finite liquid-phase concentrations, and the integral term represents the numbers of transfer units required for dilute solutions. It would be very unusual in practice to find an example in which the first (logarithmic) term is of any significance in a stripper design.

For dilute solutions in which both the operating and the equilibrium lines are straight and in which heat effects can be neglected, the integral term in Eq. (14-27) is

$$N_{OL} = \frac{1}{(1-L_M/mG_M)} \ln \left[\left(1 - \frac{L_M}{mG_M} \right) \left(\frac{x_2 - y_1/m}{x_1 - y_1/m} \right) + \frac{L_M}{mG_M} \right] \quad (14-28)$$

This equation is identical in form to Eq. (14-23). Thus, Fig. 14-5 is applicable if the concentration ratio $(x_2 - y_1/m)/(x_1 - y_1/m)$ is substituted for the abscissa and if the parameter on the curves is identified as L_M/mG_M .

Example 3: Air Stripping of VOCs from Water A 0.45-m diameter packed column was used by Dvorack et al. [*Environ. Sci. Tech.* **20**, 945 (1996)] for removing trichloroethylene (TCE) from wastewater by stripping with atmospheric air. The column was packed with 2.5-cm Pall rings, fabricated from polypropylene, to a height of 3.0 m. The TCE concentration in the entering water was 38 parts per million by weight (ppmw). A molar ratio of entering water to entering air was kept at 23.7. What degree of removal was to be expected? The temperatures of water and air were 20°C. Pressure was atmospheric.

Solution. For TCE in water, the Henry's law coefficient may be taken as 417 atm/mf at 20°C. In this low-concentration region, the coefficient is constant and equal to the slope of the equilibrium line m . The solubility of TCE in water, based on $H = 417$, is 2390 ppm. Because of this low solubility, the entire resistance to mass transfer resides in the liquid phase. Thus, Eq. (14-25) may be used to obtain N_{OL} , the number of overall liquid phase transfer units.

In the equation, the ratio $x_{BM}/(1-x)$ is unity because of the very dilute solution. It is necessary to have a value of H_L for the packing used, at given flow rates of liquid and gas. Methods for estimating H_L may be found in Sec. 5. Dvorack et al. found $H_{OL} = 0.8$ m. Then, for $h_T = 3.0$ m, $N_L = N_{OL} = 3.0/0.8 = 3.75$ transfer units.

Transfer units may be calculated from Eq. 14-25, replacing mole fractions with ppm concentrations, and since the operating and equilibrium lines are straight,

$$N_{OL} = \frac{38 - (\text{ppm})_{\text{exit}}}{\ln 38/(\text{ppm})_{\text{exit}}} = 3.75$$

Solving, $(\text{ppm})_{\text{exit}} = 0.00151$. Thus, the stripped water would contain 1.51 parts per billion of TCE.

Use of HTU and $K_G a$ Data In estimating the size of a commercial gas absorber or liquid stripper it is desirable to have data on the

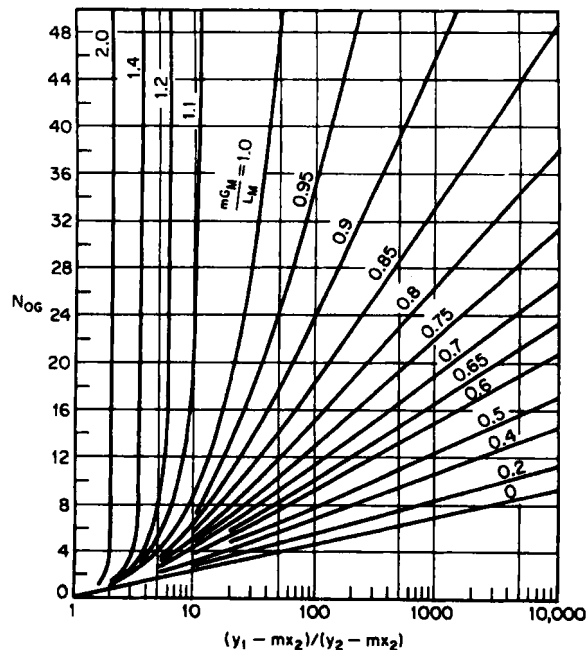


FIG. 14-5 Number of overall gas-phase mass-transfer units in a packed absorption tower for constant mG_M/L_M ; solution of Eq. (14-23). (From Sherwood and Pigford, *Absorption and Extraction*, McGraw-Hill, New York, 1952.)

overall mass transfer coefficients (or heights of transfer units) for the system of interest, and at the desired conditions of temperature, pressure, solute concentration, and fluid velocities. Such data should be obtained in an apparatus of pilot-plant or semiworks size to avoid abnormalities of scaleup. It must be remembered that values of the mass-transfer parameters are dependent not only on the phase properties and mass throughput in the contactor but also on the type of device used. Within the packing category, there are both random and ordered (structured) type packing elements. Physical characteristics of these devices will be described later.

When no $K_G a$ or HTU data are available, their values may be estimated by means of a generalized model. A summary of useful models is given in Section 5, Table 5-28. The values obtained may then be combined by the use of Eq. 14-19 to obtain values of H_{OG} and H_{OL} . This procedure is not valid, however, when the rate of absorption is limited by a chemical reaction.

Use of HETP Data for Absorber Design Distillation design methods (see Sec. 13) normally involve determination of the number of theoretical equilibrium stages or plates N . Thus, when packed towers are employed in distillation applications, it is common practice to rate the efficiency of tower packings in terms of the height of packing equivalent to one theoretical plate (HETP).

The HETP of a packed-tower section, valid for either distillation or dilute-gas absorption and stripping systems in which constant molal overflow can be assumed and in which no chemical reactions occur, is related to the height of one overall gas-phase mass-transfer unit H_{OC} by the equation

$$\text{HETP} = H_{OC} \frac{\ln(mG_M/L_M)}{(mG_M/L_M - 1)} \quad (14-29)$$

For gas-absorption systems in which the inlet gas is concentrated, the correct equation is

$$\text{HETP} = \left(\frac{y_{BM}^o}{1-y} \right)_{av} H_{OC} \frac{\ln(mG_M/L_M)}{mG_M/L_M - 1} \quad (14-30)$$

where the correction term $y_{BM}^0/(1-y)$ is averaged over each individual theoretical plate. The equilibrium compositions corresponding to each theoretical plate may be estimated by the methods described in the subsection "Plate-Tower Design." These compositions are used in conjunction with the local values of the gas and liquid flow rates and

the equilibrium slope m to obtain values for H_C , H_L , and H_{OC} corresponding to the conditions on each theoretical stage, and the local values of the HETP are then computed by Eq. (14-30). The total height of packing required for the separation is the summation of the individual HETPs computed for each theoretical stage.

PLATE-TOWER DESIGN

The design of a plate tower for gas-absorption or gas-stripping operations involves many of the same principles employed in distillation calculations, such as the determination of the number of theoretical plates needed to achieve a specified composition change (see Sec. 13). Distillation differs from gas absorption in that it involves the separation of components based on the distribution of the various substances between a gas phase and a liquid phase when all the components are present in both phases. In distillation, the new phase is generated from the original feed mixture by vaporization or condensation of the volatile components, and the separation is achieved by introducing reflux to the top of the tower.

In gas absorption, the new phase consists of an inert nonvolatile solvent (absorption) or an inert nonsoluble gas (stripping), and normally no reflux is involved. The following paragraphs discuss some of the considerations peculiar to gas-absorption calculations for plate towers and some of the approximate design methods that can be employed when simplifying assumptions are valid.

Graphical Design Procedure Construction of design diagrams (xy diagrams showing the equilibrium and operating curves) should be an integral part of any design involving the distribution of a single solute between an inert solvent and an inert gas. The number of theoretical plates can be stepped off rigorously provided the curvatures of the operating and equilibrium lines are correctly accounted for in the diagram. This procedure is valid even though an insoluble inert gas is present in the gas phase and an inert nonvolatile solvent is present in the liquid phase.

Figure 14-6 illustrates the graphical method for a three-theoretical-plate system. Note that in gas absorption the operating line is above the equilibrium curve, whereas in distillation this does not happen. In gas stripping, the operating line will be below the equilibrium curve.

On Fig. 14-6, note that the stepping-off procedure begins on the operating line. The starting point x_f, y_3 represents the compositions of the entering lean wash liquor and of the gas exiting from the top of the tower, as determined by the design specifications. After three steps one reaches the point x_1, y_f , representing the compositions of the solute-rich feed gas y_f and of the solute-rich liquor leaving the bottom of the tower x_1 .

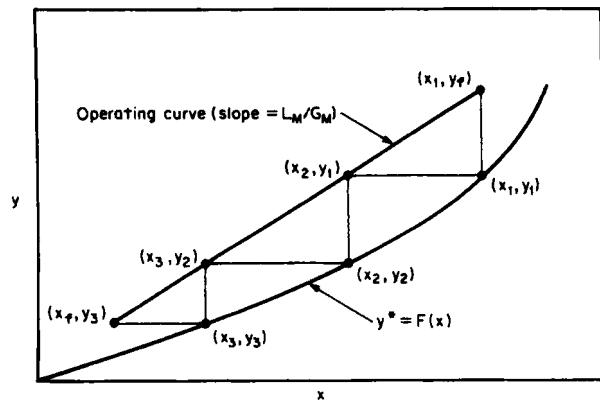


FIG. 14-6 Graphical method for a three-theoretical-plate gas-absorption tower with inlet-liquor composition x_f and inlet-gas composition y_f .

Algebraic Method for Dilute Gases By assuming that the operating and equilibrium curves are straight lines and that heat effects are negligible, Souders and Brown [*Ind. Eng. Chem.*, **24**, 519 (1932)] developed the following equation:

$$(y_1 - y_2)/(y_1 - y_2^0) = (A^{N+1} - A)/(A^{N+1} - 1) \quad (14-31)$$

where N = number of theoretical plates, y_1 = mole-fraction solute in the entering gas, y_2 = mole-fraction solute in the leaving gas, $y_2^0 = mx_2$ = mole-fraction solute in equilibrium with the incoming solvent liquor (zero for a pure solvent), and A = absorption factor = L_M/mG_M . Note that the absorption factor is the reciprocal of the expression given in Eq. (14-4) for packed columns.

When $A = 1$, Eq. (14-31) is indeterminate, and for this case the solution is given by

$$(y_1 - y_2)/(y_1 - y_2^0) = N/(N + 1) \quad (14-32)$$

Although Eq. (14-31) is convenient for computing the composition of the exit gas as a function of the number of theoretical stages, an alternative equation derived by Colburn [*Trans. Am. Inst. Chem. Eng.*, **35**, 211 (1939)] is more useful when the number of theoretical plates is the unknown:

$$N = \frac{\ln [(1 - A^{-1})(y_1 - y_2^0)/(y_2 - y_2^0) + A^{-1}]}{\ln(A)} \quad (14-33)$$

The numerical results obtained by using either Eq. (14-31) or Eq. (14-33) are identical. Thus, the two equations may be used interchangeably as the need arises.

Comparison of Eqs. (14-33) and (14-23) shows that

$$N_{OC}/N = \ln(A)/(1 - A^{-1}) \quad (14-34)$$

thus revealing the close relationship between theoretical stages in a plate tower and mass-transfer units in a packed tower. Equations (14-23) and (14-33) are related to each other by virtue of the relation

$$h_T = H_{OC}N_{OC} = (\text{HETP})N \quad (14-35)$$

Algebraic Method for Concentrated Gases When the feed gas is concentrated, the absorption factor, which is defined in general as $A = L_M/KG_M$ where $K = y^0/x$, can vary throughout the tower owing to changes in the equilibrium K values due to temperature increases. An approximate solution to this problem can be obtained by substitution of the "effective" absorption factors A_e and A' derived by Edminster [*Ind. Eng. Chem.*, **35**, 837 (1943)] into the equation

$$\frac{y_1 - y_2}{y_1} = \left[1 - \frac{1}{A'} \frac{(L_M x)_2}{(G_M y)_1} \right] \frac{A_e^{N+1} - A_e}{A_e^{N+1} - 1} \quad (14-36)$$

where subscripts 1 and 2 refer to the bottom and top of the tower respectively and the absorption factors are defined by the equations

$$A_e = \sqrt{A_1(A_2 + 1) + 0.25} - 0.5 \quad (14-37)$$

$$A' = A_1(A_2 + 1)/(A_1 + 1) \quad (14-38)$$

This procedure has been applied to the absorption of C_5 and lighter hydrocarbon vapors into a lean oil, for example.

Stripping Equations When the liquid feed is dilute and the operating and equilibrium curves are straight lines, the stripping equations analogous to Eqs. (14-31) and (14-33) are

$$(x_2 - x_1)/(x_2 - x_1^0) = (S^{N+1} - S)/(S^{N+1} - 1) \quad (14-39)$$

where $x_1^0 = y_1/m$; $S = mG_M/L_M = A^{-1}$; and

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$$N = \frac{\ln [(1-A)(x_2 - x_1^*) / (x_1 - x_1^*) + A]}{\ln(S)} \quad (14-40)$$

For systems in which the concentrations are large and the stripping factor S may vary along the tower, the following Edmister equations [Ind. Eng. Chem., **35**, 837 (1943)] are applicable:

$$\frac{x_2 - x_1}{x_2} = \left[1 - \frac{1}{S'} \frac{(G_M y)_1}{(L_M x)_2} \right] \frac{S_e^{N+1} - S_e}{S_e^{N+1} - 1} \quad (14-41)$$

where
$$S_e = \sqrt{S_2(S_1 + 1) + 0.25} - 0.5 \quad (14-42)$$

$$S' = S_2(S_1 + 1)/(S_2 + 1) \quad (14-43)$$

and the subscripts 1 and 2 refer to the bottom and top of the tower respectively.

Equations (14-37) and (14-42) represent two different ways of obtaining an effective factor, and a value of A_e obtained by taking the reciprocal of S_e from Eq. (14-42) will not check exactly with a value of A_e derived by substituting $A_1 = 1/S_1$ and $A_2 = 1/S_2$ into Eq. (14-37). Regardless of this fact, the equations generally give reasonable results for approximate design calculations.

It should be noted that throughout this section the subscripts 1 and 2 refer to the bottom and to the top of the apparatus respectively regardless of whether it is an absorber or a stripper. This has been done to maintain internal consistency among all the equations and to prevent the confusion created in some derivations in which the numbering system for an absorber is different from the numbering system for a stripper.

Tray Efficiencies in Plate Absorbers and Strippers Computations of the number of theoretical plates N assume that the liquid on each plate is completely mixed and that the vapor leaving the plate is in equilibrium with the liquid. In actual practice a condition of complete equilibrium cannot exist since interphase mass transfer requires a finite driving-force difference. This leads to the definition of an overall plate efficiency

$$E = N_{\text{theoretical}}/N_{\text{actual}} \quad (14-44)$$

which can be correlated to system design variables.

Mass-transfer theory indicates that for trays of a given design the factors most likely to influence E in absorption and stripping towers are the physical properties of the fluids and the dimensionless ratio mG_M/L_M . Systems in which the mass transfer is gas-film-controlled may be expected to have plate efficiencies as high as 50 to 100 percent, whereas plate efficiencies as low as 1 percent have been reported for the absorption of gases of low solubility (large m) into solvents of relatively high viscosity.

The fluid properties are represented by the Schmidt numbers of the gas and liquid phases. For gases, the Schmidt numbers normally are close to unity and are independent of temperature and pressure. Thus, the gas-phase mass-transfer coefficients are relatively independent of the system.

By contrast, the liquid-phase Schmidt numbers range from about 10^2 to 10^4 and depend strongly on the temperature. The effect of temperature on the liquid-phase mass-transfer coefficient is related primarily to changes in the liquid viscosity with temperature, and this derives primarily from the strong dependency of the liquid-phase Schmidt number upon viscosity.

Consideration of the preceding discussion in connection with the relationship between mass-transfer coefficients (see Sec. 5):

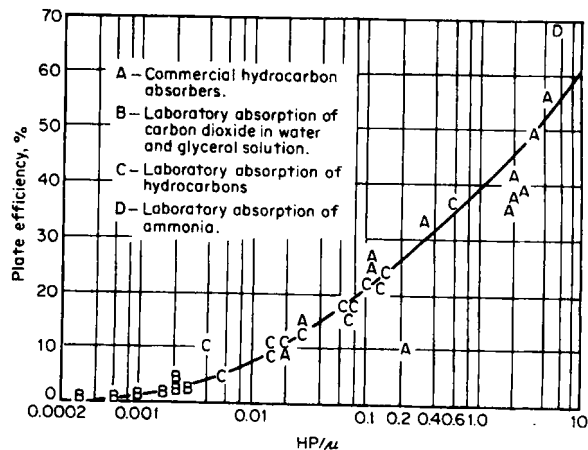


FIG. 14-7 O'Connell correlation for overall column efficiency E_{ox} for absorption. To convert HP/μ in pound-moles per cubic foot-centipoise to kilogram-moles per cubic meter-pascal-second, multiply by 1.60×10^4 . [O'Connell, Trans. Am. Inst. Chem. Eng., **42**, 741 (1946).]

$$1/K_c = (1/k_c + m/k_L) \quad (14-45)$$

indicates that variations in the overall resistance to mass transfer in absorbers and strippers are related primarily to variations in the liquid-phase viscosity μ and to variations in the slope m . A correlation of the efficiency of plate absorbers in terms of the viscosity of the liquid solvent and the solubility of the solute gas was developed by O'Connell [Trans. Am. Inst. Chem. Eng., **42**, 741 (1946)]. The O'Connell correlation for plate absorbers is presented here as Fig. 14-7.

The best procedure for making plate-efficiency corrections (which obviously can be quite large) is to use experimental-test data from a prototype system that is large enough to be representative of an actual commercial tower.

Example 4: Actual Plates for Steam Stripping The number of actual plates required for steam-stripping an acetone-rich liquor containing 0.573 mole percent acetone in water is to be estimated. The design overhead recovery of acetone is 99.9 percent, leaving 18.5 ppm weight of acetone in the stripper bottoms. The design operating temperature and pressure are 101.3 kPa and 94°C respectively, the average liquid-phase viscosity is 0.30 cP, and the average value of $K = y^*/x$ for these conditions is 33.

By choosing a value of $mC_M/L_M = S = A^{-1} = 1.4$ and noting that the stripping medium is pure steam (i.e., $x_1^* = 0$), the number of theoretical trays according to Eq. (14-40) is

$$N = \frac{\ln [(1 - 0.714)(1000) + 0.714]}{\ln(1.4)} = 16.8$$

The O'Connell parameter for gas absorbers is $\rho_L/KM\mu_L$, where ρ_L is the liquid density, lb/ft³; μ_L is the liquid viscosity, cP; M is the molecular weight of the liquid; and $K = y^*/x$. For the present design

$$\rho_L/KM\mu_L = 60.1/(33 \times 18 \times 0.30) = 0.337$$

and according to the O'Connell graph for absorbers (Fig. 14-7) the overall tray efficiency for this case is estimated to be 30 percent. Thus, the required number of actual trays is $16.8/0.3 = 56$ trays.

HEAT EFFECTS IN GAS ABSORPTION

Overview One of the most important considerations involved in designing gas-absorption towers is to determine whether or not temperatures will vary along the length of the tower because of heat effects, since the solubility of the solute gas normally depends strongly upon the temperature. When heat effects can be neglected, computation of the tower dimensions and required flows is relatively straight-

forward, as indicated by the simplified design methods discussed earlier for both packed and plate absorbers and strippers. When heat effects cannot be neglected, the computational problem becomes much more difficult.

Heat effects that may cause temperatures to vary from point to point in a gas absorber are (1) the heat of solution of the solute