SEPARATION OPERATIONS 1 (0905451)

06 – ABSORPTION AND STRIPPING

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Outline

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- Phase Contacting Equipment
  - Tray Columns
  - Packed Columns
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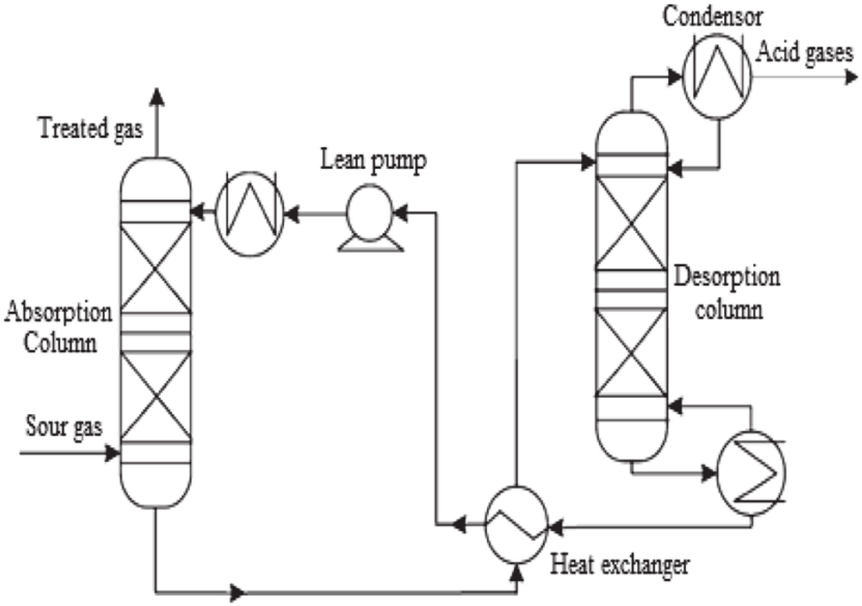
- Kremser Group Method for Multicomponent Absorption or Stripping
- Rate Based Methods for Packed Columns

Please see https://encyclopedia.che.engin.umich.edu/absorbers/

Definitions

- Absorption, or scrubbing, is the removal of a component (the solute or absorbate) from a gas stream via a (ideally) nonvolatile liquid (the solvent or absorbent).
- Desorption, or stripping, is the removal of a component from a liquid stream via vaporization and uptake by an (ideally) insoluble gas stream.
- Absorption and stripping are opposite unit operations, and are often used together as a cycle.
- Both absorption and stripping can be operated as equilibrium stage processes using trayed columns or, more commonly, using packed columns.



Natural gas purification based on amine absorption-striping cycle

- The amount of each component absorbed depends on the number of equilibrium stages and the component’s absorption factor. $A_i = \dfrac{L}{K_i V}$

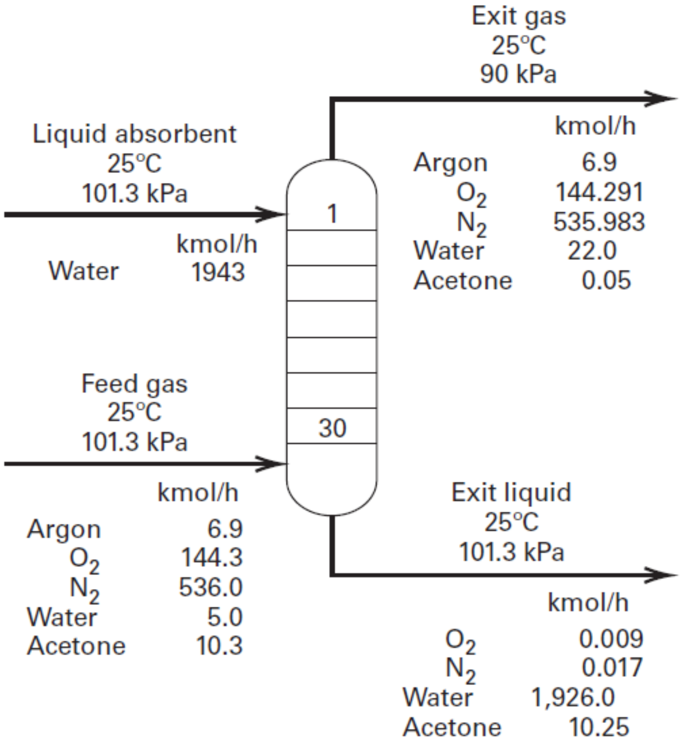


Figure 6.1 Industrial absorption process.

Component	$\mathcal{A}_i = L/K_iV$	K-value
Water	89.2	0.031
Acetone	1.38	2.0
Oxygen	0.00006	45,000
Nitrogen	0.00003	90,000
Argon	0.00008	35,000

Industrial Applications

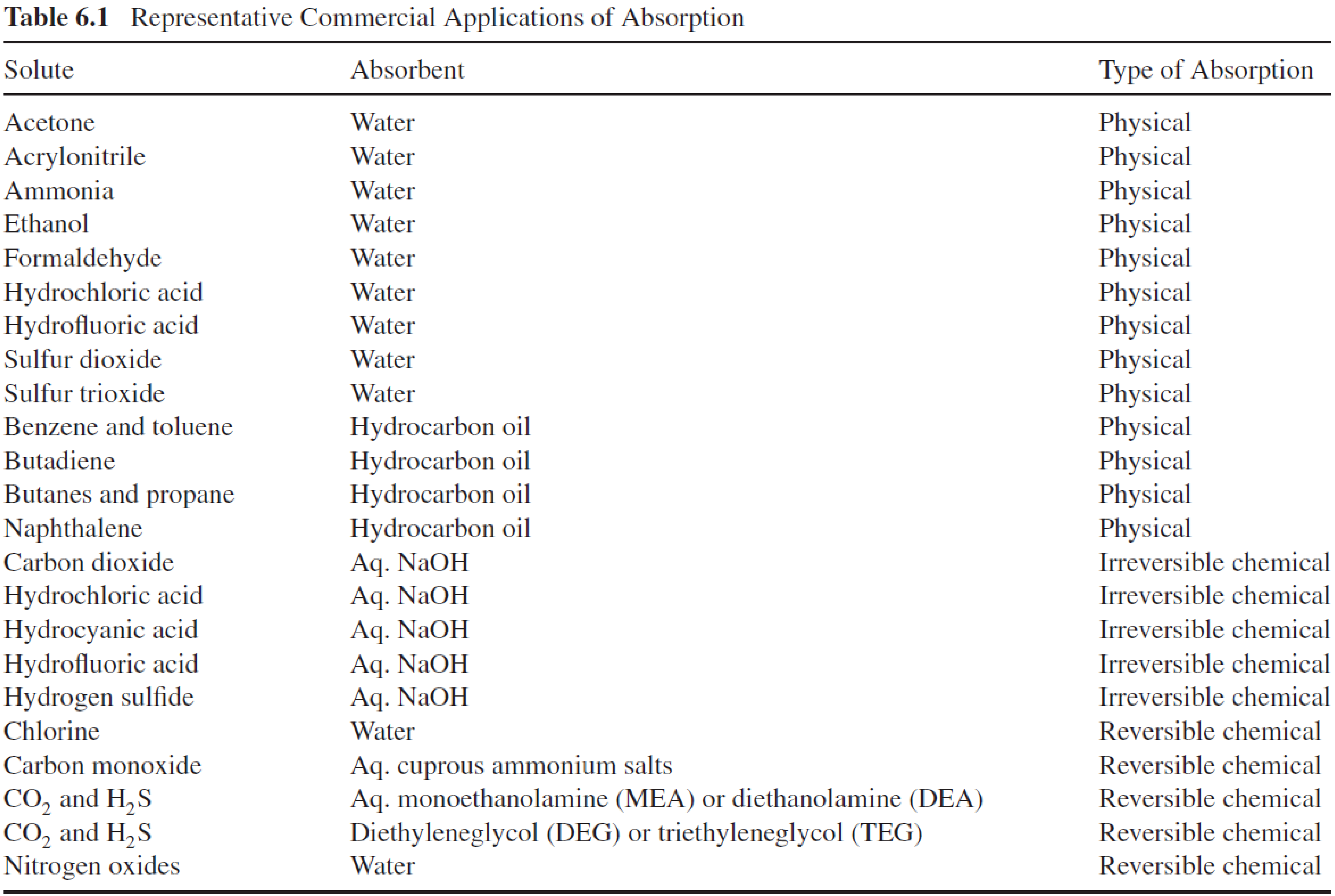
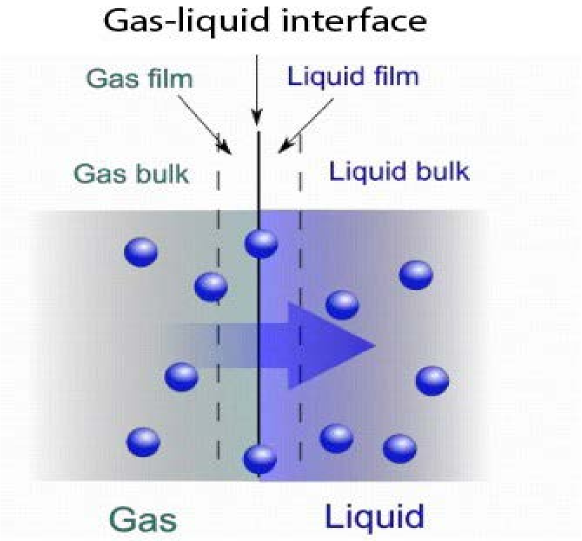


Table 6.1 Representative Commercial Applications of Absorption

Solute	Absorbent	Type of Absorption
Acetone	Water	Physical
Acrylonitrile	Water	Physical
Ammonia	Water	Physical
Ethanol	Water	Physical
Formaldehyde	Water	Physical
Hydrochloric acid	Water	Physical
Hydrofluoric acid	Water	Physical
Sulfur dioxide	Water	Physical
Sulfur trioxide	Water	Physical
Benzene and toluene	Hydrocarbon oil	Physical
Butadiene	Hydrocarbon oil	Physical
Butanes and propane	Hydrocarbon oil	Physical
Naphthalene	Hydrocarbon oil	Physical
Carbon dioxide	Aq. NaOH	Irreversible chemical
Hydrochloric acid	Aq. NaOH	Irreversible chemical
Hydrocyanic acid	Aq. NaOH	Irreversible chemical
Hydrofluoric acid	Aq. NaOH	Irreversible chemical
Hydrogen sulfide	Aq. NaOH	Irreversible chemical
Chlorine	Water	Reversible chemical
Carbon monoxide	Aq. cuprous ammonium salts	Reversible chemical
CO_2 and H_2S	Aq. monoethanolamine (MEA) or diethanolamine (DEA)	Reversible chemical
CO_2 and H_2S	Diethyleneglycol (DEG) or triethyleneglycol (TEG)	Reversible chemical
Nitrogen oxides	Water	Reversible chemical

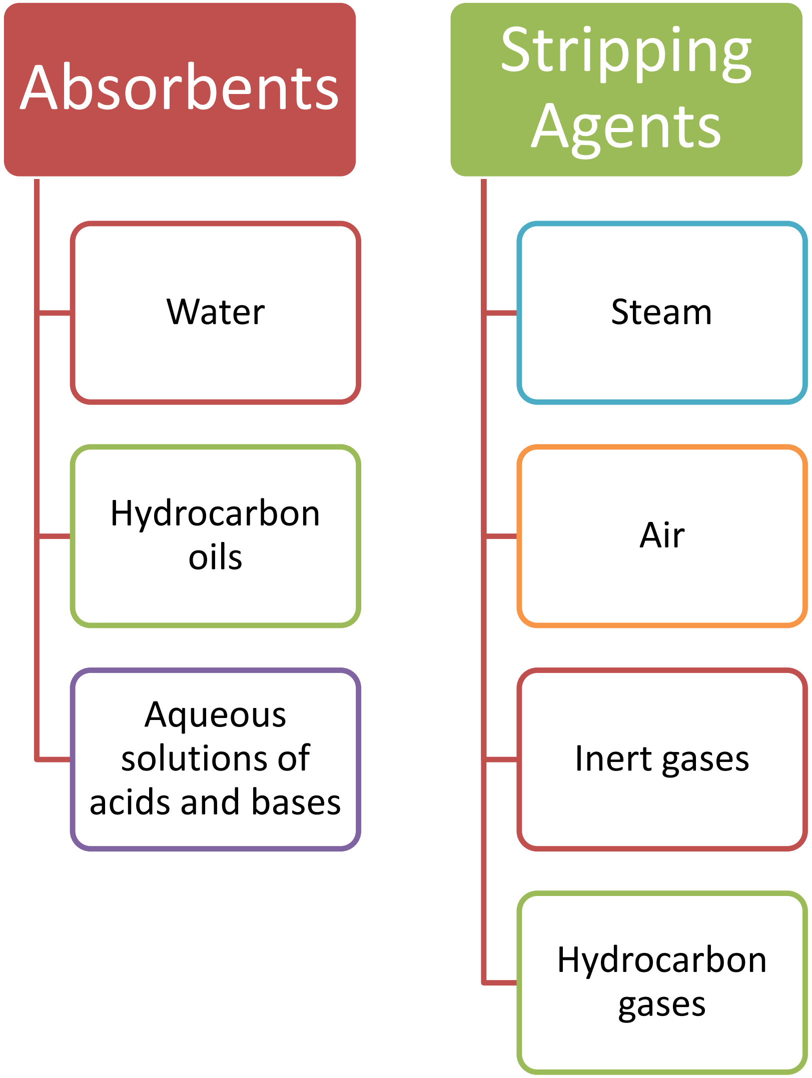
The Ideal Absorbent

- High solubility for solute(s).
- Low volatility to reduce loss.
- Stability and inertness;
- Low corrosiveness.
- Low viscosity and high diffusivity.
- Low foaming proclivities;
- Low toxicity and flammability.



Absorption

Widely Used Absorbents and Stripping Agents



Absorbents
- Water
- Hydrocarbon oils
- Aqueous solutions of acids and bases

Stripping Agents
- Steam
- Air
- Inert gases
- Hydrocarbon gases

Issues of Setting *P* and *T*

- Absorber operating pressure should be high and temperature low in order to minimize stage requirements and/or absorbent flow rate, and to lower the equipment volume required to accommodate the gas flow.
  - Unfortunately, both compression and refrigeration of a gas are expensive. Therefore, most absorbers are operated at feed-gas pressure, which may be greater than ambient pressure, and at ambient temperature, which can be achieved by cooling the feed gas and absorbent with cooling water, unless one or both streams already exist at an ambient or subambient temperature.
- Operating pressure should be low and temperature high for a stripper to minimize stage requirements and stripping agent flow rate.
  - However, because the maintenance of a vacuum is expensive, and steam jet exhausts are polluting, strippers are commonly operated at a pressure just above ambient.
  - A high temperature can be used, but it should not be so high as to cause vaporization or undesirable chemical reactions. The possibility of phase changes occurring can be checked by bubble-point and dew-point calculations.

Phase Contacting Equipment

- When multiple stages are required, phase contacting is mostly carried out in columns:
  - The most common devices are cylindrical, vertical columns containing trays or packing.
  - Less common are spray towers, bubble columns, and centrifugal contactors.

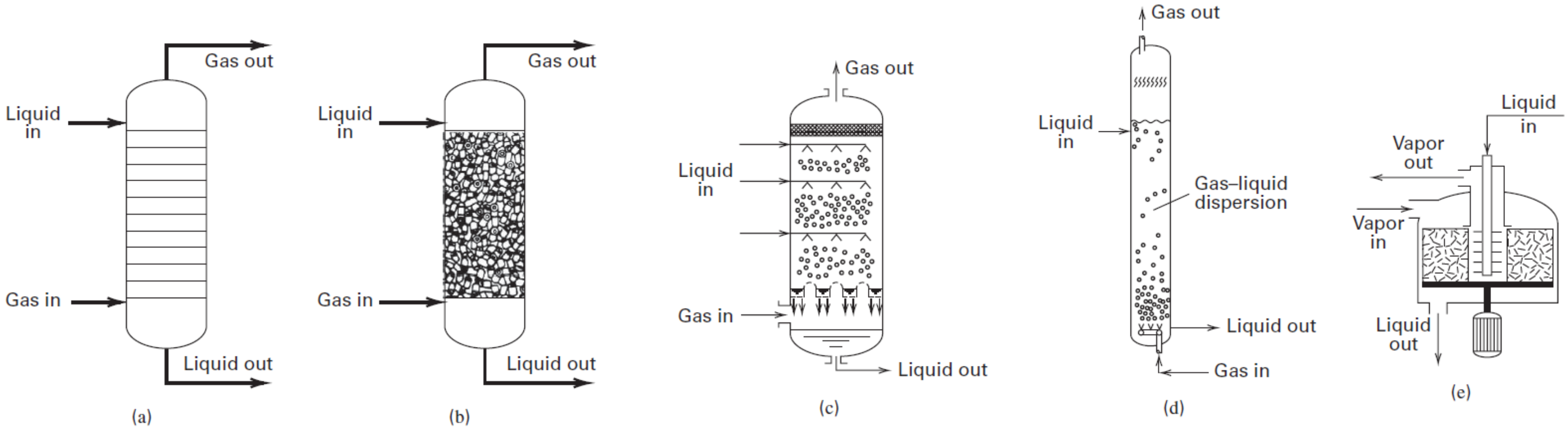


Figure 6.2 Industrial equipment for absorption and stripping: (a) trayed tower; (b) packed column; (c) spray tower; (d) bubble column; (e) centrifugal contactor.

Tray Columns

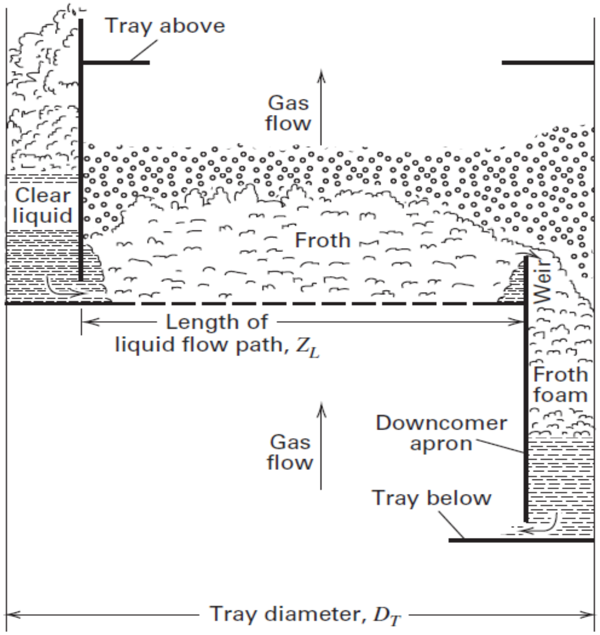


Figure 6.3 Vapor–liquid contact on a tray. [Adapted from B.F. Smith, *Design of Equilibrium Stage Processes*, McGraw-Hill, New York (1963).]

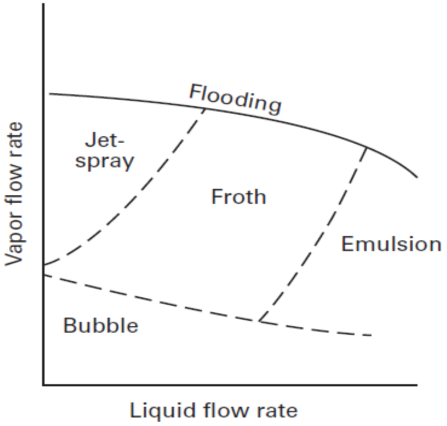


Figure 6.4 Possible vapor–liquid flow regimes for a contacting tray.

Table 6.2 Comparison of Tray Types

	Sieve Trays	Valve Trays	Bubble-Cap Trays
Relative cost	1.0	1.2	2.0
Pressure drop	Lowest	Intermediate	Highest
Efficiency	Lowest	Highest	Highest
Vapor capacity	Highest	Highest	Lowest
Typical turndown ratio	2	4	5

Operational flexibility is reported in terms of turndown ratio (ratio of maximum-to-minimum vapor flow capacity).

Packed Columns

- A packed column is a vertical, cylindrical vessel containing one or more sections of packing over whose surface the liquid flows downward as a film on the packing and walls, or as droplets between packing elements.
- Feed gas enters at the bottom, passes through a vapor distributor, and flows upward through the wetted packing, thus contacting the liquid and passing out the top.
- Liquid enters at the top into a liquid distributor, flows downward through the packed sections, and leaves at the bottom.

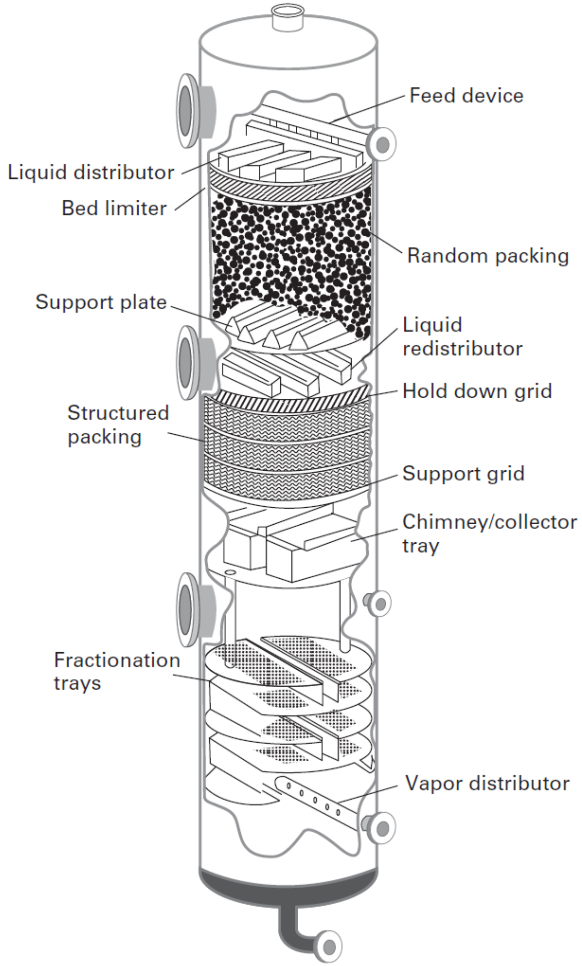


Figure 6.6 Details of internals in a column with packing and trays.

Types of Packing

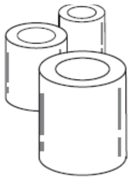
Packing

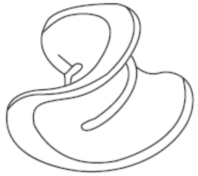
Random (dumped) packing.

Structured packing of crimped layers of mesh or corrugated sheets.

Grid packing that have an open-lattice structure.

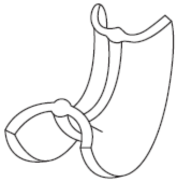
- Mainly for heat transfer and washing.


Ceramic Raschig rings


Ceramic Berl saddle

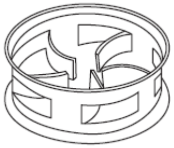

Sulzer® Wire Gauze

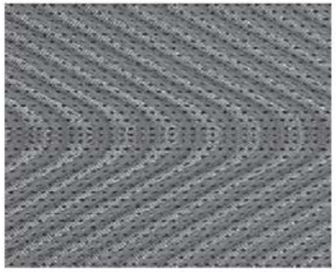

Metal Pall® ring


Ceramic Intalox® saddle


Mellapak®


Metal Intalox® IMTP


Metal Cascade Mini-Ring®(CMR)


MellapakPlus®


Metal Raschig Super-Ring®


Metal Intalox® Ultra™

https://www.sulzer.com/en/shared/products/column-internals
https://koch-glitsch.com/

Table 6.3 Comparison of Types of Packing

	Random		
	Raschig Rings and Saddles	"Through Flow"	Structured
Relative cost	Low	Moderate	High
Pressure drop	Moderate	Low	Very low
Efficiency	Moderate	High	Very high
Vapor capacity	Fairly high	High	High
Typical turndown ratio	2	2	2

Choice Between Trays and Packing

- Trayed towers can be designed more reliably.
- Use of structured packing should be avoided at pressures above 200 psi and liquid flow rates above 10 gpm/ft^2.
- Structured packing, though expensive, are the best choice for installations when pressure drop is a factor or for replacing existing trays (**retrofitting**) when a higher capacity or degree of separation is required.
- Turbulent liquid flow is desirable if mass transfer is limiting in the liquid phase, and a continuous, turbulent gas flow is desirable if mass transfer is limiting in the gas phase.
    - Usually, the (continuous) gas phase is mass-transfer-limiting in packed columns and the (continuous) liquid phase is mass-transfer-limiting in tray columns.

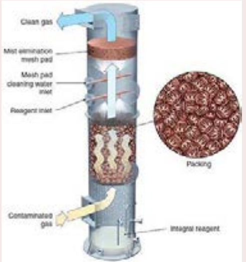
Choice Between Trays and Packing

Pros

- Low pressure drop required e.g., vacuum systems.
- Small diameters possible (< 0.5 m).
- Can handle foaming systems.
- Low capital, operating, and maintenance cost.
- Simple construction.
- Can handle corrosive materials due to corrosion-resistant packing.
- Can handle severe foaming systems.
- Reduces back mixing in comparison to spray columns.
- Better mass transfer than in spray columns

Cons

- Fewer stages compared to other columns.
- Channeling, which must be controlled by redistributing liquid.
- Cannot handle extremely high or low flow rates.
- Cannot handle liquids with high viscosities.
- Need to be preferentially wetted to avoid reduction of the interfacial area to volume ratio.



Pros

- The liquid/vapor contact in the cross-flow of plate columns is more effective than the countercurrent flow in packed columns.
- Can handle high or low liquid flow rates cost-effectively
- Can handle solids.
- Easily customized to specific requirements such as operations requiring much heat.

Cons

- Higher pressure drops than packed columns.
- Slow reaction rate processes.
- Plugging and fouling may occur.

Graphical Methods for Tray Towers

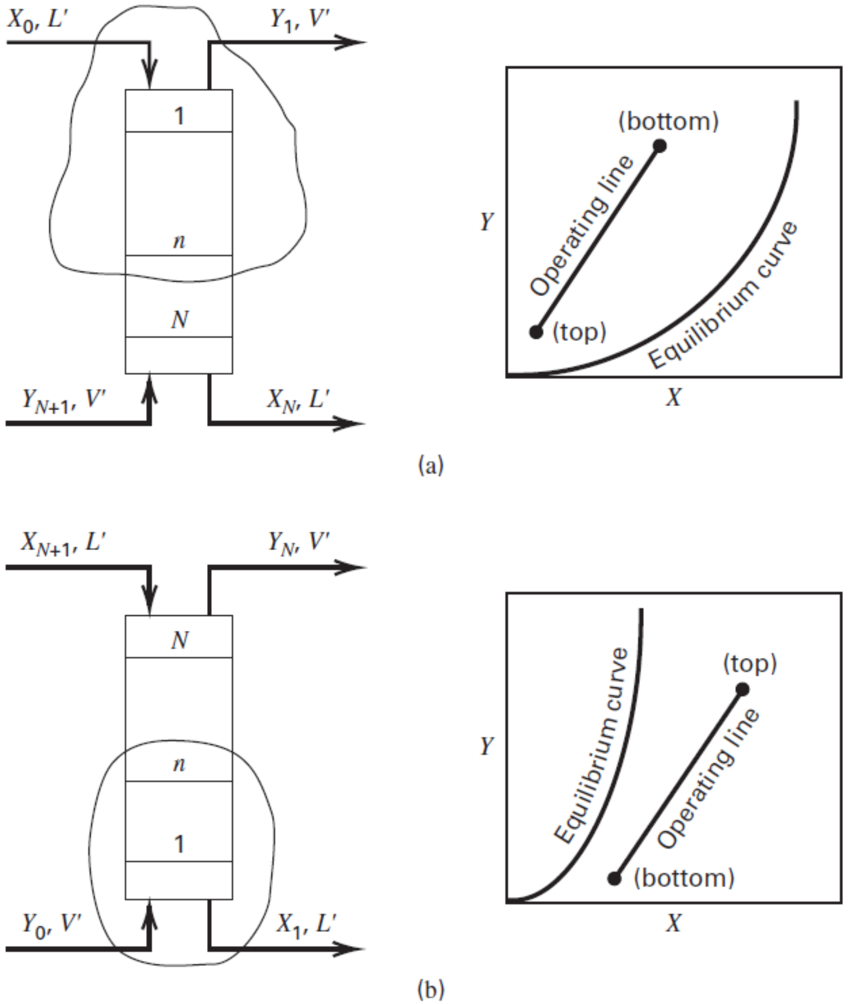


Figure 6.9 Continuous, steady-state operation in a countercurrent cascade with equilibrium stages: (a) absorber; (b) stripper.

L' = molar flow rate of solute-free absorbent

V' = molar flow rate of solute-free gas (carrier gas)

X = mole ratio of solute to solute-free absorbent in the liquid

Y = mole ratio of solute to solute-free gas in the vapor

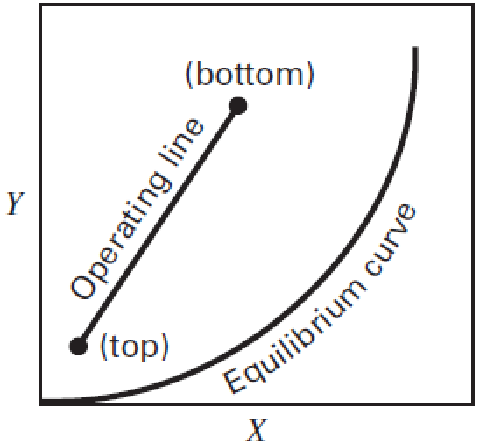
Values of *L'* and V' remain constant throughout the tower because only solute undergoes mass transfer between phases

Equilibrium Curves for the Solute

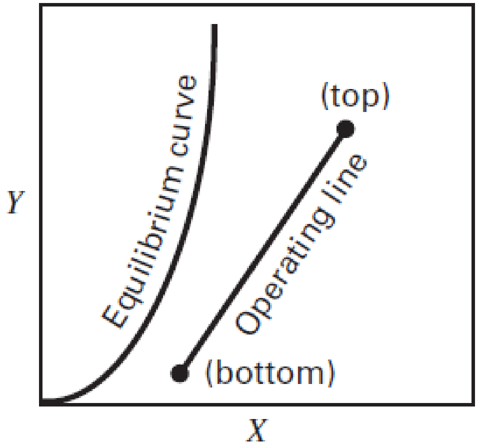
- For the solute at any stage *n*, the *K*-value is given by:

$$K_n = \frac{y_n}{x_n} = \frac{Y_n / (1+Y_n)}{X_n / (1+X_n)}$$

$$Y = \frac{y}{1-y} \quad \text{and } X = \frac{x}{1-x}$$

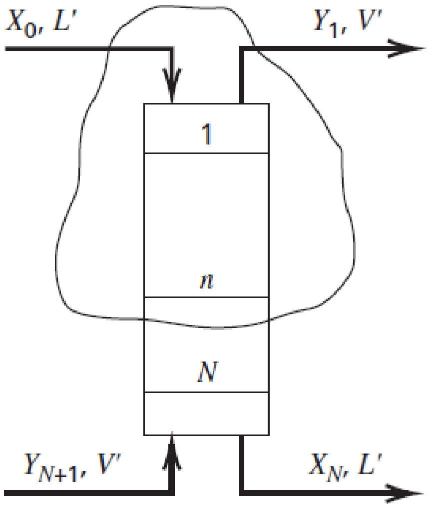


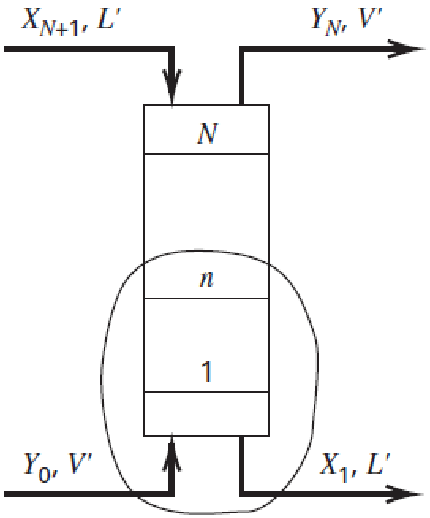
- The equilibrium curve will not be a straight line, but it will pass through the origin.
- If the solute undergoes an irreversible liquid-phase chemical reaction with the solvent to make a nonvolatile product, the equilibrium curve will be a straight line of zero slope, passing through the origin.
- For a pure stripping agent, the operating line extends to Y = 0.
- For a pure absorbent, it goes through X = 0.



Operating Lines (from Solute Material Balances)

- Exiting and entering solute compositions and solute-free flowrates are paired:
  - Absorbers
    - Pairs at the top are: (X_0, L') and (Y_1, V').
    - Pairs at the bottom are: (X_N, L') and (Y_{N+1}, V').
  - Strippers
    - Pairs at the top are: (X_{N+1}, L') and (Y_N, V').
    - Pairs at the bottom are: (X_1, L') and (Y_0, V').
- Solute material balances are written around one end of the tower and an arbitrary equilibrium stage, *n*.





Absorbers and Strippers Operating Lines

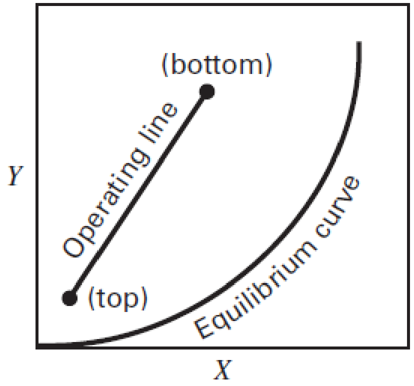
- Absorber

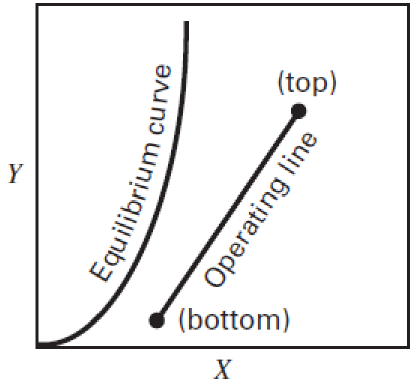
$$Y_{n+1} = X_n \frac{L'}{V'} + Y_1 - X_0 \frac{L'}{V'}$$

- Stripper

$$Y_n = X_{n+1} \frac{L'}{V'} + Y_0 - X_1 \frac{L'}{V'}$$

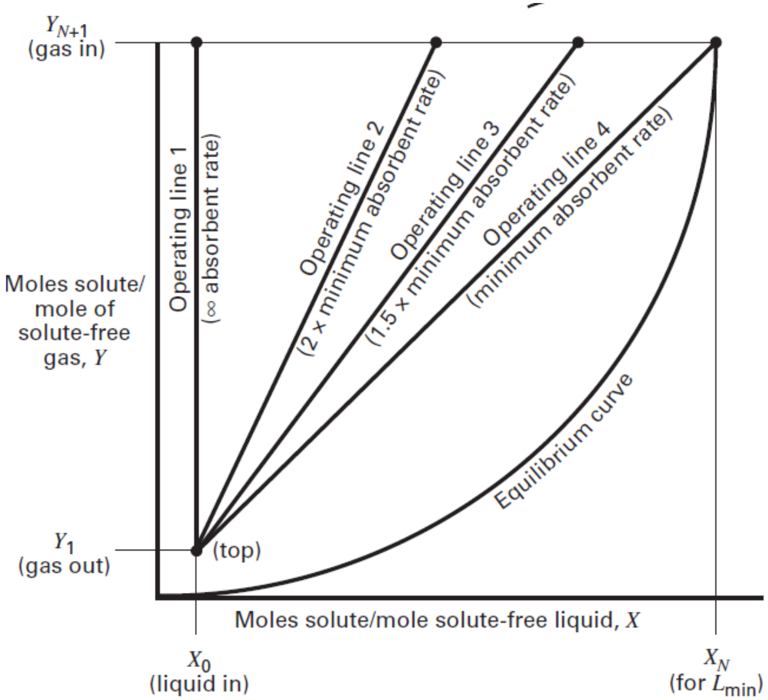
- **Operating lines** plot terminal points represent conditions at the top and bottom of the tower.
  - For absorbers, the operating line is above the equilibrium curve because, for a given solute concentration, *X*, in the liquid, the solute concentration, *Y*, in the gas is always greater than the equilibrium value, thus providing a mass-transfer driving force for absorption.
  - For strippers, operating lines lie below equilibrium lines, thus enabling desorption. Operating lines are straight with a slope of *L′/V′*.

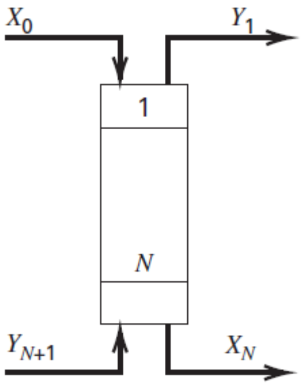




Minimum Absorbent Flow Rate (for ∞ Stages)

- Operating lines for four different solute-free absorbent flow rates, L', are shown for a fixed solute-free gas feed rate, V'.
- In each case, the solute concentration in the exiting gas, Y_1, is the same.
    - Each operating line passes through the terminal point, (Y_1, X_0), at the top of the column.
    - To achieve the desired value of Y_1 for given Y_{N+1}, X_0, and V', the solute-free absorbent flow rate L' must be between an ∞ absorbent flow with $L'/V' = \infty$, as represented by operating line 1, and a minimum absorbent rate (corresponding to ∞ stages), L'_{min}, as represented by operating line 4, with the equilibrium curve and operating line intersecting at Y_{N+1}.
- Intermediate operating lines 2 and 3, correspond to 2 and 1.5 times L'_{min}, respectively. The solute concentration in the outlet liquid, X_N, depends on L'.





- Operating line can terminate at the equilibrium line as in operating line 4 but cannot cross it because that would be a violation of the second law of thermodynamics.
- The minimum absorbent flow rate, L'_{min}, corresponds to a value of X_N (leaving the bottom of the tower) in equilibrium with Y_{N+1}, the solute concentration in the feed gas.
    - An infinite number of stages for this equilibrium to be achieved is required.

$$X_0 L' + Y_{N+1} V' = X_N L' + Y_1 V'$$

$$L' = \frac{V'\left(Y_{N+1} - Y_1\right)}{\left(X_N - X_0\right)}$$

Solute material balance over the entire absorber with *n* = *N* :

$$K_N = \frac{Y_{N+1} \big/ \left(1 + Y_{N+1}\right)}{X_N \big/ \left(1 + X_N\right)}$$

Distribution coefficient for stage *N* for the minimum absorbent rate

Combine the material balance and definition of distribution coefficient

$$L'_{\min} = \frac{V'\left(Y_{N+1} - Y_1\right)}{\left\{Y_{N+1} \big/ \left[Y_{N+1}\left(K_N - 1\right) + K_N\right]\right\} - X_0}$$

Special Cases

- For dilute solutes, where $Y \approx y$ and $X \approx x$, the minimum liquid rate approaches:

$$L'_{\min} = V' \left(\frac{y_{N+1} - y_1}{y_{N+1}/K_N - x_0} \right)$$

- For the entering liquid with $X_0 \approx 0$:

$$L'_{\min} = V' \times K_N \times \left(1 - \frac{y_1}{y_{N+1}} \right)$$

fraction of solute absorbed

- This equation confirms that:

$$L'_{\min} = \begin{cases} \uparrow \text{ with } V' \\ \uparrow \text{ with } K_N \\ \uparrow \text{ with fraction of solute absorbed} \end{cases}$$

- Absorbent flowrate is some multiple of $L'_{\min}$, typically from 1.1 - 2.

Determining Minimum Vapor Rates for Strippers

- For dilute solutes, where $Y \approx y$ and $X \approx x$, the minimum vapor rate approaches:

$$V'_{\min} = \frac{L'}{K_N} \times \left(1 - \frac{x_1}{x_{N+1}}\right)$$

fraction of solute stripped

- This equation confirms that:

$$V'_{\min} = \begin{cases} \uparrow \text{ with L}' \\ \downarrow \text{ with } K_N \\ \uparrow \text{ with fraction of solute stripped} \end{cases}$$

- Stripping factors are optimal at about 1.4.

$$S = K\frac{V}{L} = \frac{1}{A}$$

Determination of the Number of equilibrium Stages: Absorption

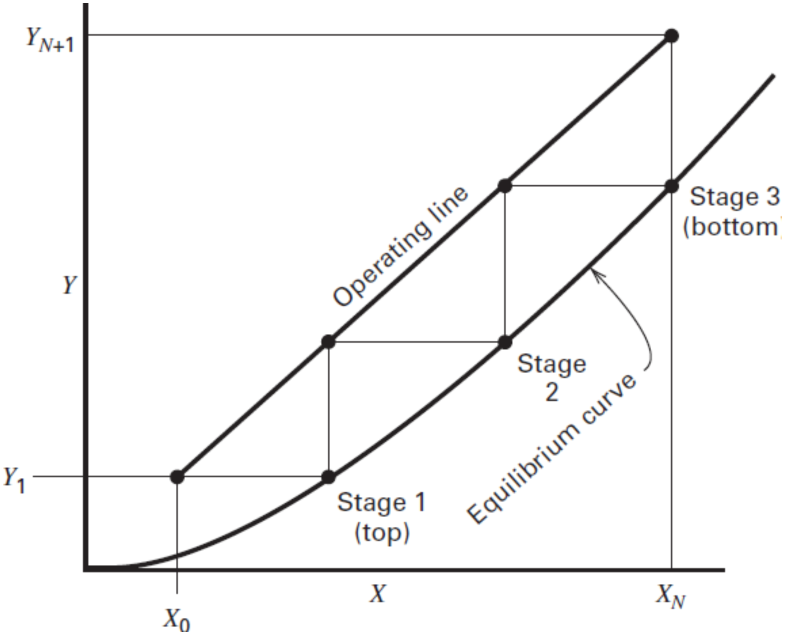
Obtain VLE curve

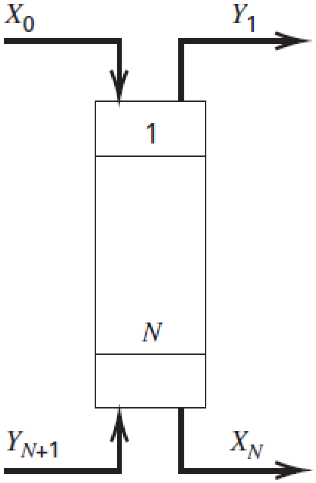
Obtain the absorption operating line (atop of the equilibrium curve for absorption).

The operating line is produced using the two points (X_0, Y_1) and (X_N, Y_{N+1}).

Step horizontally from the point (X_0, Y_1) and follow the staircase.

Count the number of stages by how many times the staircase hits the equilibrium curve.





An exact integer number of stages is rare; usually fractions of stages arise.

Determination of the Number of equilibrium Stages: Stripping

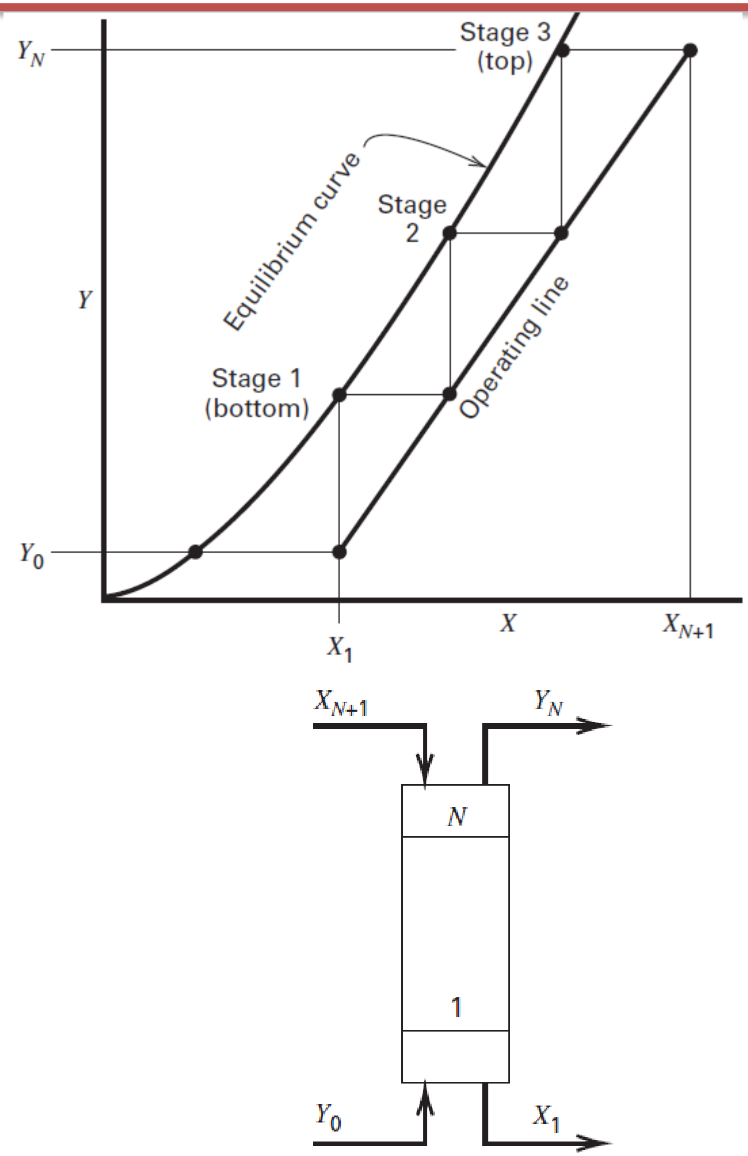
Obtain VLE curve

Obtain the stripping operating line (below the equilibrium curve for stripping).

The operating line is produced using the two points (X_1, Y_0) and (X_{N+1}, Y_N).

Step horizontally from the point (X_{N+1}, Y_N) and follow the staircase.

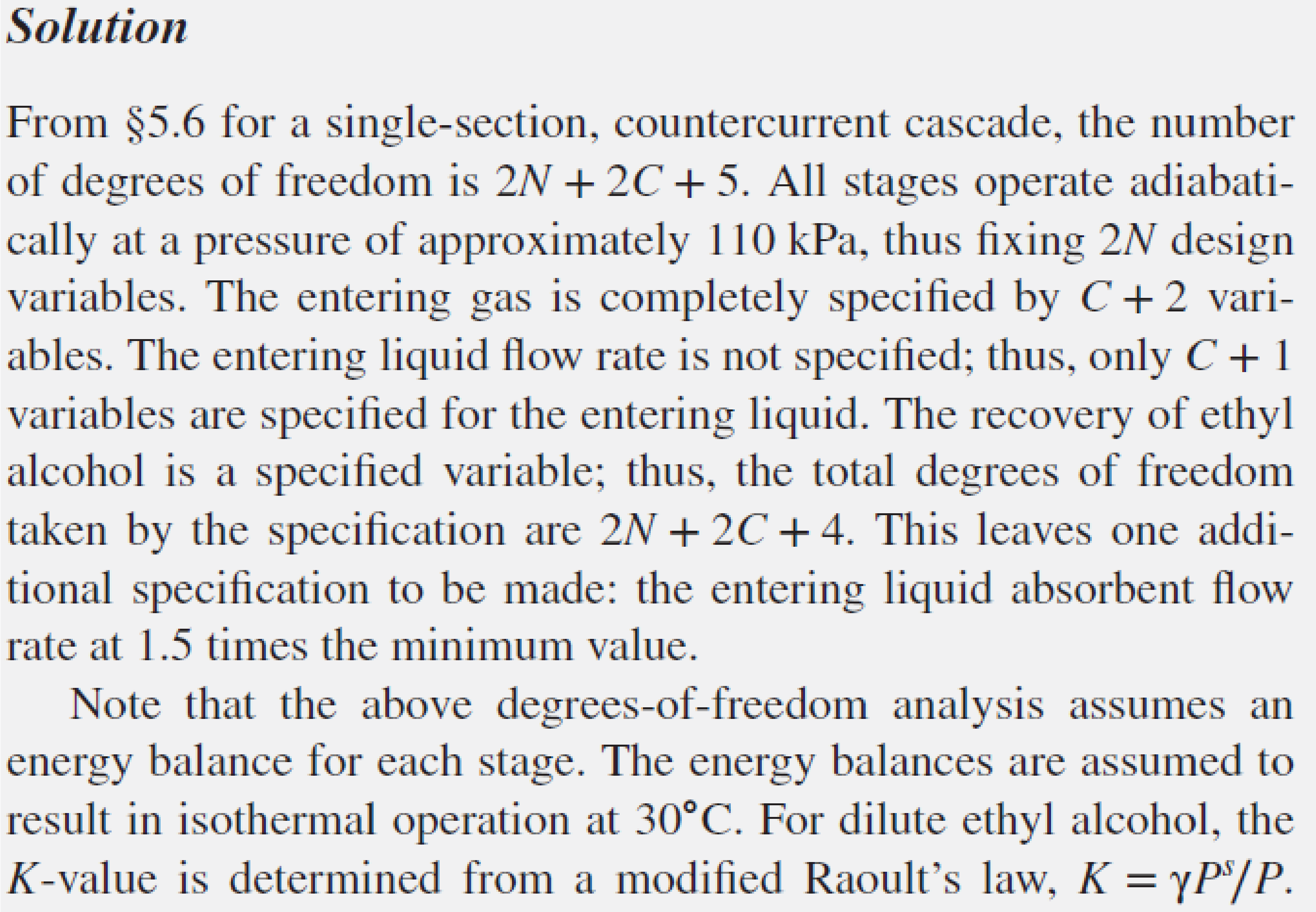
Count the number of stages by how many times the staircase hits the equilibrium curve.



An exact integer number of stages is rare; usually fractions of stages arise.

EXAMPLE 6.1 Recovery of Ethyl Alcohol.

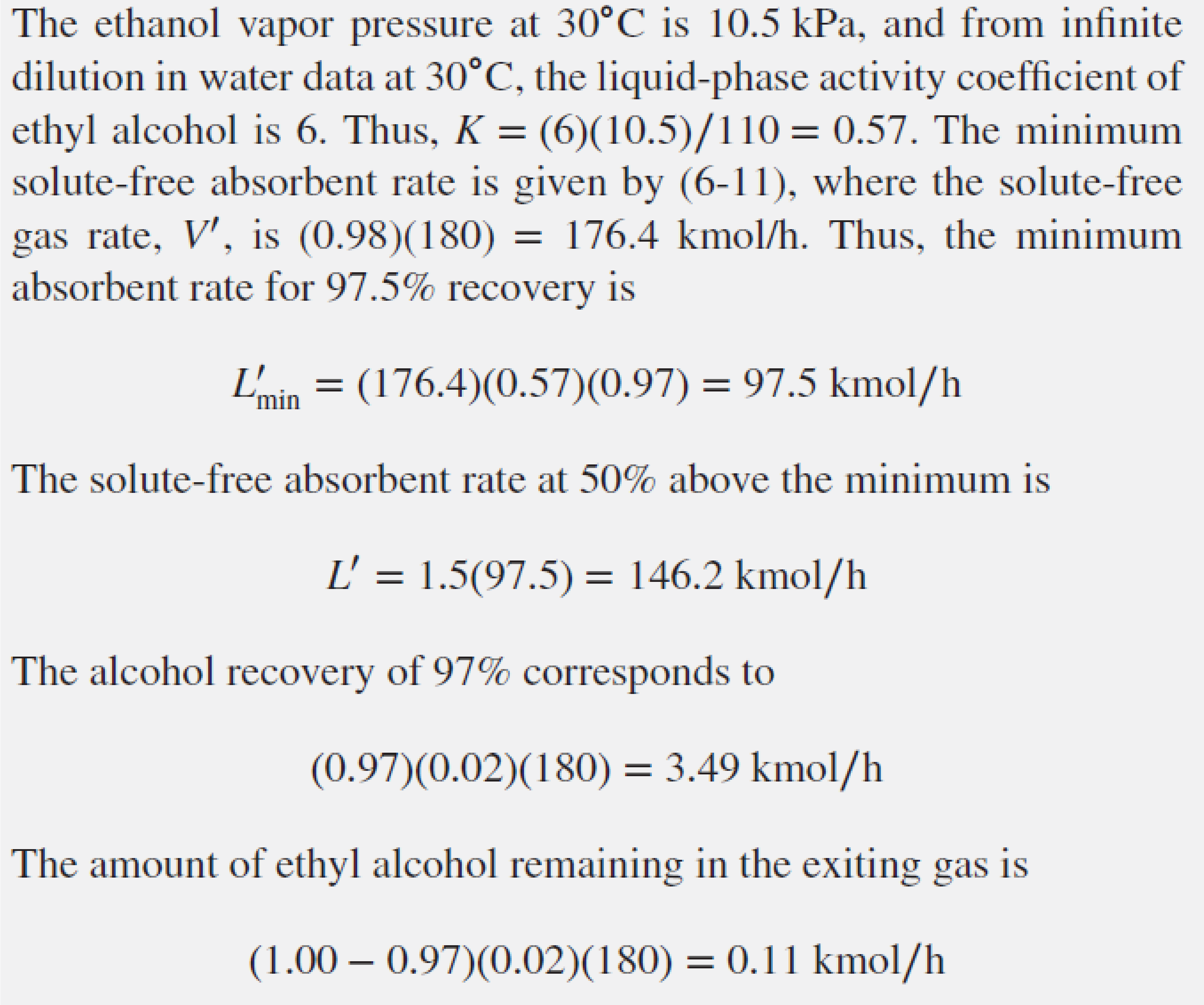
In a bioprocess, molasses is fermented to produce a liquor containing ethyl alcohol. A CO_2-rich vapor with a small amount of ethyl alcohol is evolved. The alcohol is recovered by absorption with water in a sieve-tray tower. Determine the number of equilibrium stages required for countercurrent flow of liquid and gas, assuming isothermal, isobaric conditions and absorption of ethanol only. Entering gas is 180 kmol/h, 98% CO_2 and 2% ethyl alcohol, 30°C, 110 kPa. Entering liquid absorbent is 100% water, 30°C, 110 kPa. Required recovery (absorption) of ethyl alcohol is 97%.



Solution

From §5.6 for a single-section, countercurrent cascade, the number of degrees of freedom is $2N + 2C + 5$. All stages operate adiabatically at a pressure of approximately 110 kPa, thus fixing $2N$ design variables. The entering gas is completely specified by $C + 2$ variables. The entering liquid flow rate is not specified; thus, only $C + 1$ variables are specified for the entering liquid. The recovery of ethyl alcohol is a specified variable; thus, the total degrees of freedom taken by the specification are $2N + 2C + 4$. This leaves one additional specification to be made: the entering liquid absorbent flow rate at 1.5 times the minimum value.

Note that the above degrees-of-freedom analysis assumes an energy balance for each stage. The energy balances are assumed to result in isothermal operation at 30°C. For dilute ethyl alcohol, the K-value is determined from a modified Raoult's law, $K = \gamma P^s/P$.



The ethanol vapor pressure at 30°C is 10.5 kPa, and from infinite dilution in water data at 30°C, the liquid-phase activity coefficient of ethyl alcohol is 6. Thus, $K = (6)(10.5)/110 = 0.57$. The minimum solute-free absorbent rate is given by (6-11), where the solute-free gas rate, V', is (0.98)(180) = 176.4 kmol/h. Thus, the minimum absorbent rate for 97.5% recovery is

$$L'_{\text{min}} = (176.4)(0.57)(0.97) = 97.5 \text{ kmol/h}$$

The solute-free absorbent rate at 50% above the minimum is

$$L' = 1.5(97.5) = 146.2 \text{ kmol/h}$$

The alcohol recovery of 97% corresponds to

$$(0.97)(0.02)(180) = 3.49 \text{ kmol/h}$$

The amount of ethyl alcohol remaining in the exiting gas is

$$(1.00 - 0.97)(0.02)(180) = 0.11 \text{ kmol/h}$$

Alcohol mole ratios at both ends of the operating line are as follows:

$$\text{top of the column}\left\{ X_0 = 0,\ \ Y_1 = \frac{0.11}{176.4} = 0.0006 \right\}$$

$$\text{bottom of the column}\left\{ Y_{N+1} = \frac{0.11 + 3.49}{176.4} = 0.0204, \right.$$

$$\left. X_N = \frac{3.49}{146.2} = 0.0239 \right\}$$

The equation for the operating line from (6-3), with $X_0 = 0$, is

$$Y_{N+1} = \left(\frac{146.2}{176.4}\right) X_N + 0.0006 = 0.829 X_N + 0.0006 \qquad (1)$$

This is a dilute system. From (6-1), the equilibrium curve, using $K = 0.57$, is

$$0.57 = \frac{Y/(1+Y)}{X/(1+X)}$$

Solving for Y,

$$Y = \frac{0.57X}{1 + 0.43X} \qquad (2)$$

For the coordinates to cover the entire column, the necessary range of X for a plot of Y versus X is 0 to almost 0.025. From the Y–X equation, (2), the following values are obtained:

Y	X
0.00000	0.000
0.00284	0.005
0.00569	0.010
0.00850	0.015
0.01130	0.020
0.01410	0.025

For this dilute ethyl alcohol system, the maximum error in Y is 1.0%, if Y is taken simply as $Y = KX = 0.57X$. The equilibrium curve in Figure 6.13 is plotted using the Y–X data. It is almost a straight line. The operating line drawn through the terminal points (Y_1, X_0) and (Y_{N+1}, X_N) is straight. The equilibrium stages are stepped off as shown, starting at the top stage (Y_1, X_0) located in the plot near the lower left corner. The required number of equilibrium stages, N, for 97% absorption of ethyl alcohol is between 6 and 7, at about 6.1.

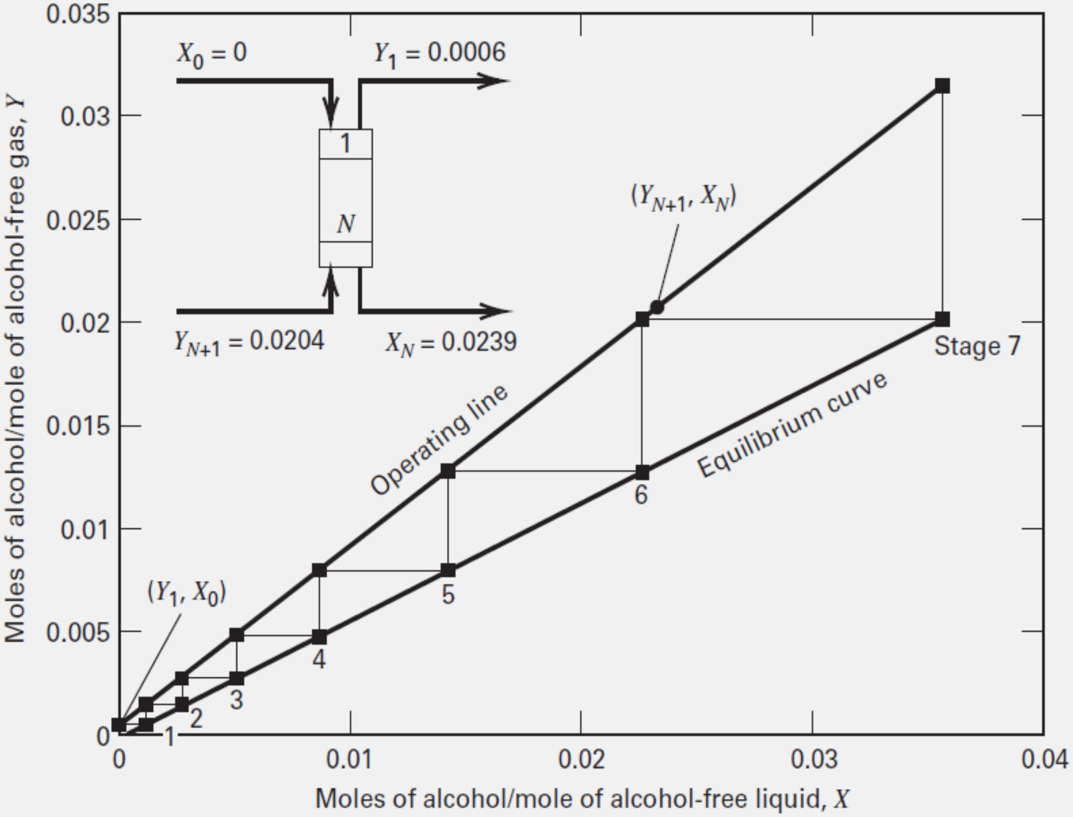


Figure 6.13 Graphical determination of number of equilibrium stages for an absorber.

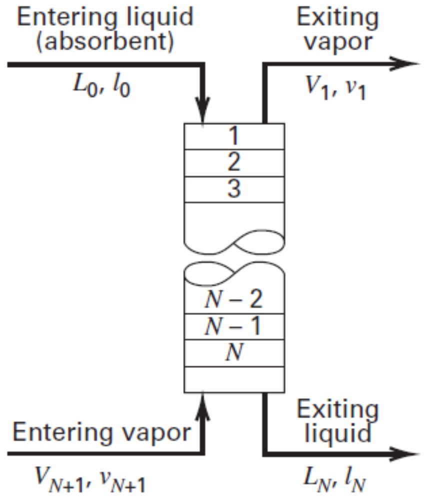
Kremser Group Method for Multicomponent Absorption or Stripping

- Graphical method becomes unsuitable for:
  1. Problem specifications fix the number of stages rather than the percent recovery of solute.
  2. More than one solute is absorbed or stripped
  3. The location of the equilibrium curve is unknown because the optimal operating T and P have not been established
  4. Very low or very high concentrations of solute force the construction to the corners of the diagram so that multiple Y–X diagrams of varying scales are needed to achieve accuracy.
- When the graphical method is unsuitable, the analytical method of Kremser is useful for obtaining results that can be used to initialize rigorous methods in process simulators.
  - Group method that relates the number of equilibrium stages to the recovery of one key component in a single-section, countercurrent cascade used for multicomponent absorption or stripping.
  - Called a **group method** because it only provides an overall treatment of the group of stages in the cascade.
  - The procedure does not consider detailed changes in T, P, phase compositions, and flow rates from stage to stage.

Kremser Equations for Absorption and Stripping

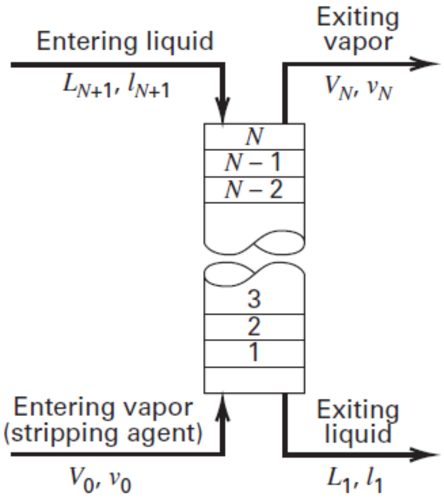
- Absorption: fraction of species not absorbed

$$\phi_A = \frac{A_e - 1}{A_e^{N+1} - 1} \qquad A = \frac{L}{KV}$$



- Stripping: fraction of species not stripped

$$\phi_S = \frac{S_e - 1}{S_e^{N+1} - 1} \qquad S = K\frac{V}{L}$$



$$(N+1) = \frac{\ln\left(\dfrac{A_e - 1}{\phi_A} + 1\right)}{\ln A_e}$$

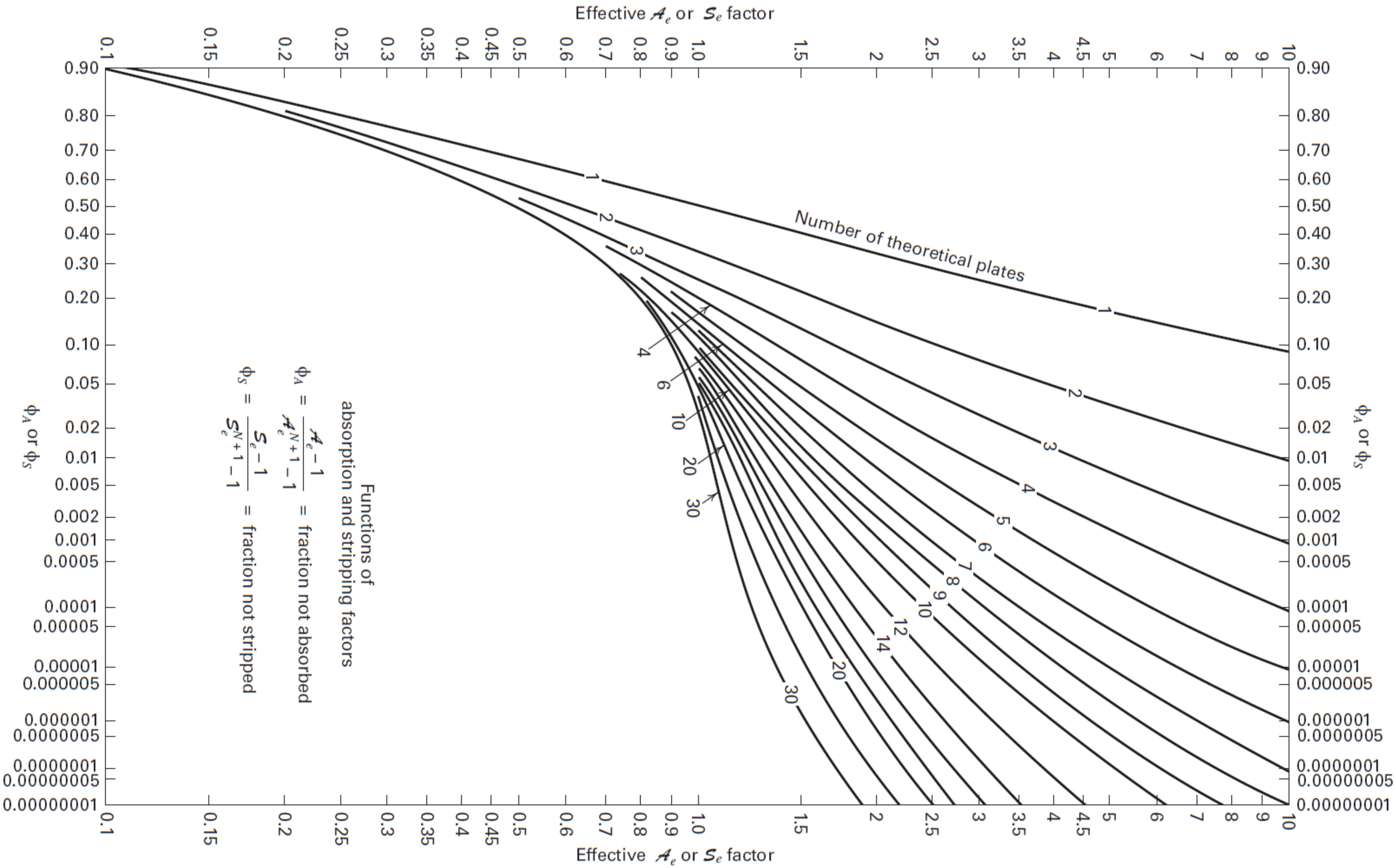


Figure 6.15 Plot of the Kremser equation for a single-section, countercurrent cascade.

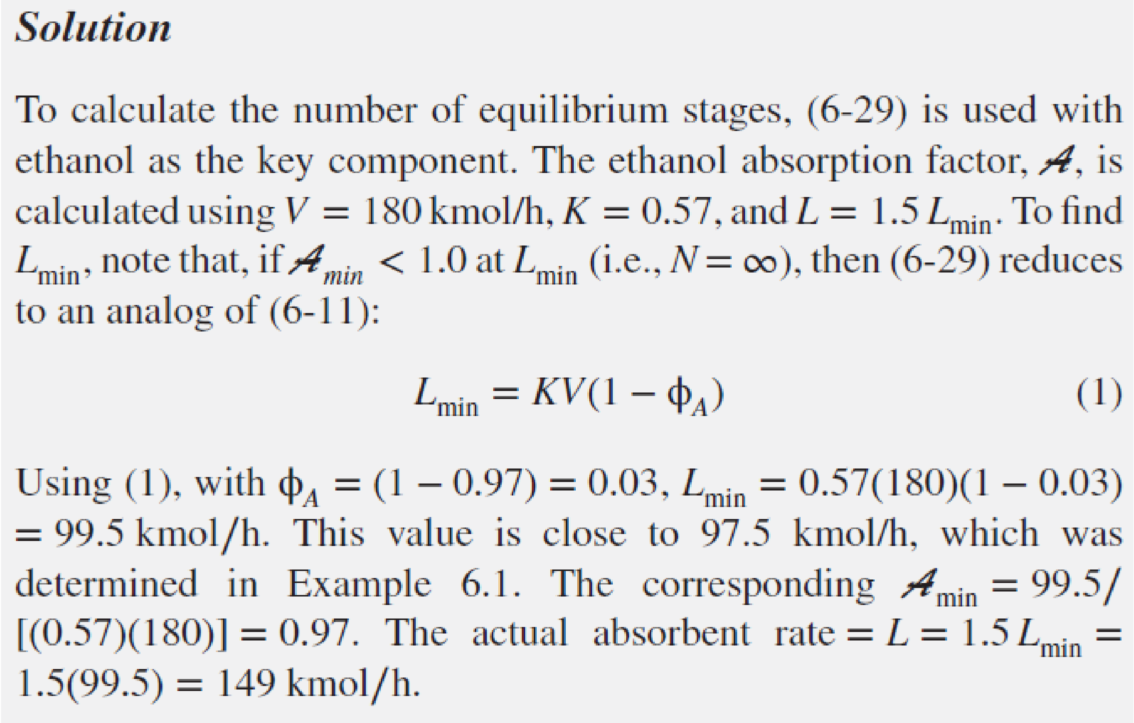
EXAMPLE 6.2 Recovery of Ethyl Alcohol Using the Kremser Equation.

Repeat Example 6.1 for the recovery of ethyl alcohol from CO_2 by absorption in water, using the Kremser method. Assume the absorber operates at 30ºC and 110 kPa. The total entering gas flow rate is 180 kmol/h with 2 mol% ethanol. The absorbent is pure water. The specified recovery of ethanol is 97%. From Example 6.1, the *K*-value for ethanol is 0.57. Assuming an entering absorbent flow rate of 1.5 times the minimum, determine the following:

(a) The number of equilibrium stages required

(b) The fraction of the entering water stripped into the gas

(c) The fraction of the entering CO_2 absorbed by the water



Solution

To calculate the number of equilibrium stages, (6-29) is used with ethanol as the key component. The ethanol absorption factor, $\mathcal{A}$, is calculated using $V = 180$ kmol/h, $K = 0.57$, and $L = 1.5\,L_{\min}$. To find $L_{\min}$, note that, if $\mathcal{A}_{min} < 1.0$ at $L_{\min}$ (i.e., $N = \infty$), then (6-29) reduces to an analog of (6-11):

$$L_{\min} = KV(1 - \phi_A) \tag{1}$$

Using (1), with $\phi_A = (1 - 0.97) = 0.03$, $L_{\min} = 0.57(180)(1 - 0.03) = 99.5$ kmol/h. This value is close to 97.5 kmol/h, which was determined in Example 6.1. The corresponding $\mathcal{A}_{\min} = 99.5/[(0.57)(180)] = 0.97$. The actual absorbent rate $= L = 1.5\,L_{\min} = 1.5(99.5) = 149$ kmol/h.

(a) To find N from (6-29), compute $\mathcal{A}_e = 149/[(0.57)(180)] = 1.45$, which is close to the optimal value. From (6-29),

$$\phi_A = 0.03 = \frac{\mathcal{A}_e - 1}{\mathcal{A}_e^{N+1} - 1} = \frac{1.45 - 1}{1.45^{N+1} - 1} \qquad (2)$$

Equation (2) is a nonlinear equation in N. Using MATLAB with fsolve, $N = 6.46$. This compares to slightly more than 6 stages, as determined graphically in Example 6.1

(b) For water, use a Raoult's law K-value with a water vapor pressure at 30°C = 4.24 kPa. Therefore, from (6-37), $K = 4.24/110 = 0.0385$. From (6-32), $\mathcal{S}_e$ for water $= 0.0385(180)/149 = 0.0465$. From (6-31), with $N = 6.46$, using ethanol as the key component, the fraction of water not stripped $= \dfrac{\mathcal{S}_e - 1}{\mathcal{S}_e^{N+1} - 1} = \dfrac{0.0385 - 1}{0.0385^{6.45+1} - 1} = 0.962$, and the fraction of entering water stripped $= 1 - 0.962 = 0.038$.

(c) For CO_2 absorption into water, use the Henry's law K-value from (6-39), with $H_{CO_2} = 1900$ atm $= 195000$ kPa from Figure 4.18. From (6-39), K for $CO_2 = 295000/110 = 1770$, and from (6-18), $\mathcal{A}_e = 149/[1770(180)] = 0.00047$. From (6-29),

$$\frac{\mathcal{A}_e - 1}{\mathcal{A}_e^{N+1} - 1} = \frac{0.00047 - 1}{0.00047^{6.45+1} - 1} = 0.9995$$

$$= \phi_A = \text{fraction of } CO_2 \text{ not absorbed}$$

Therefore, the fraction of CO_2 absorbed $= 1 - 0.9995 = 0.0005$.

Rate Based Methods for Packed Columns

- Packed columns are continuous, differential-contacting devices that do not have physically distinguishable, discrete stages.

Model	Discrete stage-wise model (Tray columns)	Differential continuous contacting model
Equilibrium between phases (VLE)	Equilibrium stage model (most common in practice!) Such as McCabe-Thiele or dilute: Fenske, Kremser	Can't be used. May replace it with HETP model.
Non equilibrium (rate based)	Assume Nonequilibrium between bulk phases. May use two film theory or surface renewal/penetration theories.	Two-film approach is the most common: For nondilute systems numerical integration is required. For dilute systems: log-mean formula is used.

Rate Based Methods for Packed Columns

- Packed columns are better described by mass transfer models than by equilibrium-stage concepts.
- In practice, packed-tower performance is often presented on the basis of equivalent equilibrium stages using a packed height equivalent to a theoretical plate/stage, the HETP or HETS:

$$\text{HETP} = \frac{\text{Packing height}}{\text{Number of equivalen equilibrium stages}} = \frac{l_T}{N_t}$$

- No theoretical basis!

Usually, the lower the HETP, the more expensive and complex the packing.

EXAMPLE HETP

To illustrate the HETP concept, consider Example 6.1, which involves the recovery of ethyl alcohol from a CO_2-rich vapor by absorption with water.

From Example 6.2, the required N_t is 6.46. If experience shows that the use of 1.5-inch metal Pall rings will produce an average HETP of 2.25 ft, then the packed height from $l_T = (\text{HETP})N_t = 2.25(6.46) = 13.7$ ft.

If metal Intalox IMTP #40 random packing has an HETP = 2.0 ft, then $l_T = 12.9$ ft.

With Mellapak 250Y sheet-metal structured packing, the HETP might be 1.2 ft, giving lT = 7.75 ft.

Kister Estimates of HETP

KISTER, H.Z., Distillation Design, McGraw-Hill, New York, (1992).

Modern random packing with low-viscosity liquids:

- HETP (ft) = 1.5xD_p (in)

Structured packing at low-to-moderate pressures and low-viscosity liquids:

- HETP(ft) = 100/*a* (ft^2/ft^3) + 0.333

Absorption with a viscous liquid

- HETP (ft) = 5 to 6

For Vacuum service:

- HETP (ft) = 1.5xD_p (in)+ 0.5

For small-diameter towers less than 2 ft in diameter:

- HETP(ft) = diameter in feet; but not less than 1 ft

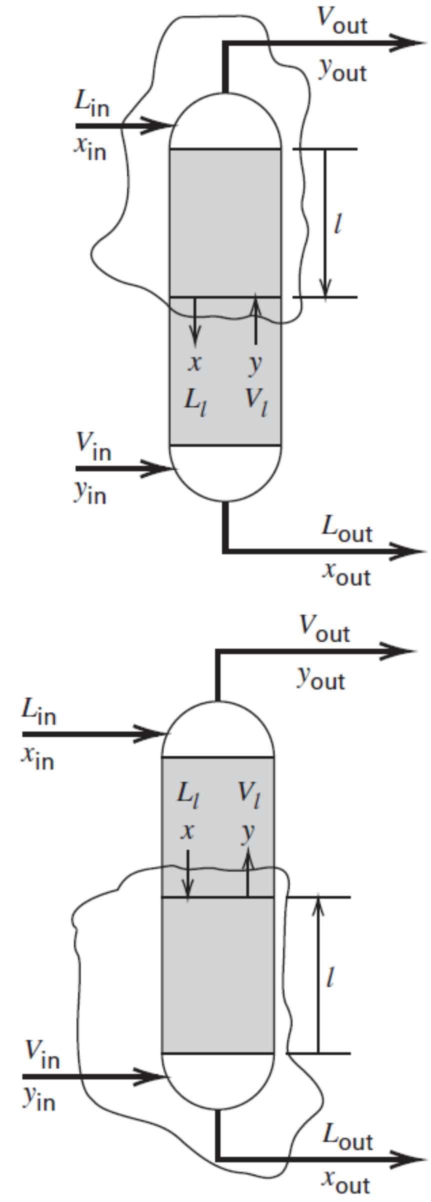
For high-pressure service with structured packing:

- HETP(ft) > 100/*a* (ft^2/ft^3) + 0.333

D_p (in) is the nominal diameter of random packing
a (ft^2/ft^3) is the specific surface area of structured packing

Mass Transfer Rate-Based Column Height

- It is preferable to determine the packed height from theoretically based methods involving mass-transfer coefficients.
- Consider the countercurrent-flow packed columns of packed height l_T shown.
- For packed absorbers and strippers, operating-line equations analogous to those in tray towers can be derived.



Deriving the Operating Line: Absorbers and Strippers

- A molar material balance for the solute, around the upper envelope of packed height, *l*, gives:

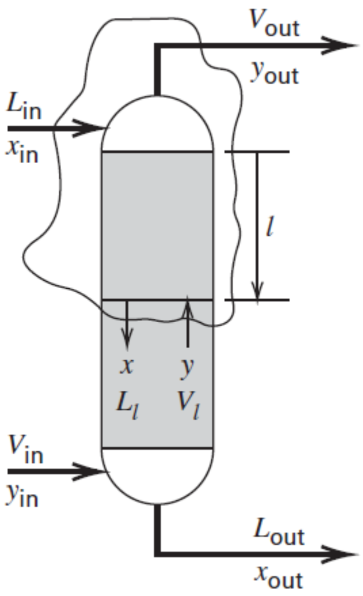
$$x_{in}L_{in} + yV_l = xL_l + y_{out}V_{out}$$

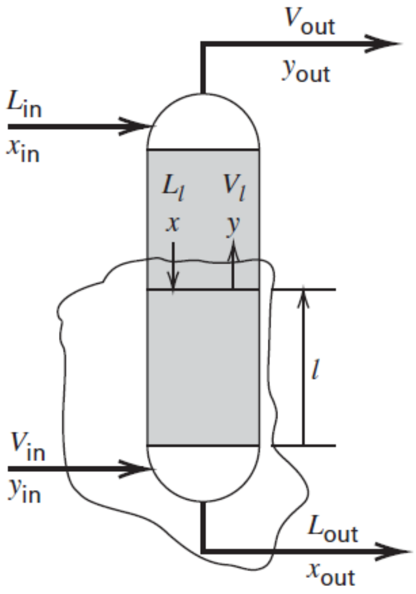
- Assuming dilute solutions such that:
  - $V_l = V_{in} = V_{out} = V$ and $L_l = L_{in} = L_{out} = L$.

$$y = x\frac{L}{V} + y_{out} - x_{in}\frac{L}{V}$$

- Similarly for a stripper

$$y = x\frac{L}{V} + y_{in} - x_{out}\frac{L}{V}$$





Two-Film Theory

- mole fractions *y* and *x* represent **bulk compositions** of the gas and liquid in contact at any vertical location in the packing.
- For absorption, with solute mass transfer from the gas to the liquid stream, the two-film theory applies.
  - A concentration gradient exists in each thin film.
  - At the **interface** between the two phases, physical equilibrium exists.
  - Thus, as with trayed towers, an operating line and an equilibrium line are of great importance for packed towers.

For a given problem specification, the location of the operating and equilibrium lines is independent of whether the tower is trayed or packed.
Determining the minimum absorbent liquid or stripping vapor flowrates in a packed column is identical to that for trayed towers.

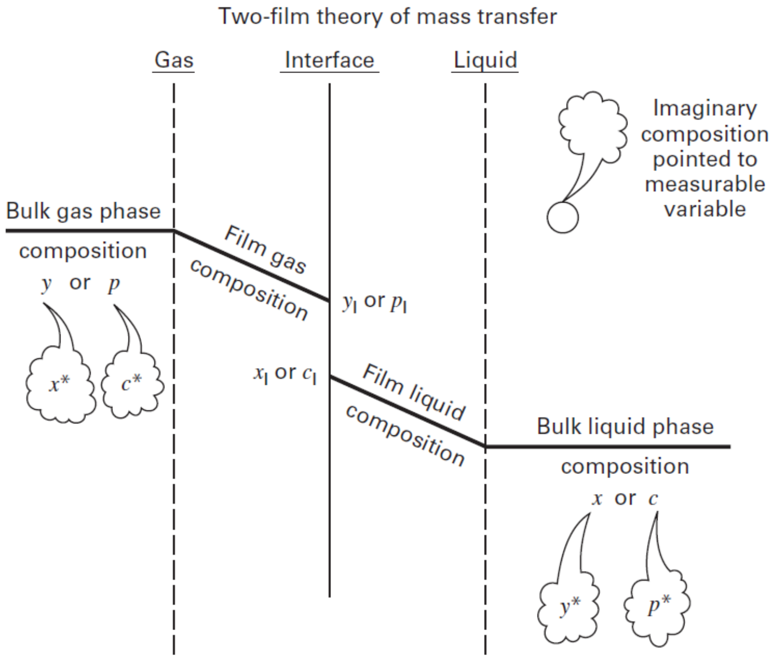


Figure 6.30 Interface properties in terms of bulk properties.

Use of Mass Transfer Coefficients and Interfacial Area

- The rate of mass transfer for absorption or stripping can be expressed in terms of mass-transfer coefficients for each phase.
  - MT Coefficients, k, based on a unit area for mass transfer could be used, but the area for interfacial mass transfer in a packed bed is difficult to determine.
  - Accordingly, as with mass transfer in the froth of a trayed tower, it is common to use **volumetric mass-transfer coefficients**, ka, where the variable a represents the interfacial area for mass transfer between phases per unit volume of packed bed.

Mass Transfer Rates

- At steady state, in the absence of chemical reactions, the rate of solute mass transfer through the gas-phase film must equal the rate through the liquid film.
  - If the system is dilute in solute, unimolecular diffusion (UMD) is approximated by the equations for equimolar counter-diffusion (EMD).
  - The solute mass-transfer rate per unit volume of packed bed, *r*, is written in terms of mole-fraction driving forces in each phase (or in terms of a partial-pressure driving force in the gas phase and a concentration driving force in the liquid).
- Using mole fractions for absorption, with the subscript I to denote the phase interface, the solute mass-transfer rate is

$$r = \underbrace{k_y a}_{\text{volumetric mass-transfer coefficient for the gas phase}} \left(y - y_I \right) = \underbrace{k_x a}_{\text{volumetric mass-transfer coefficient for the liquid phase}} \left(x_I - x \right)$$

volumetric mass-transfer coefficient for the gas phase

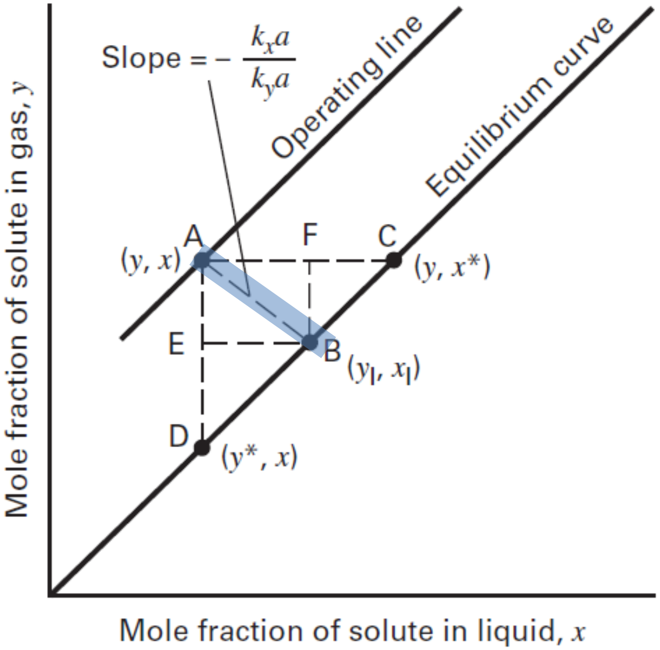
volumetric mass-transfer coefficient for the liquid phase

Interfacial Compositions

- At the interface, phase equilibrium exists between y_I and x_I.
- The composition at the interface depends on the ratio k_xa/k_ya from the definition of r:

$$\frac{y - y_I}{x - x_I} = -\frac{k_x a}{k_y a}$$

- A straight line of slope $-k_xa/k_ya$, drawn from the operating line at the bulk composition point (y, x), intersects the equilibrium curve at (y_I, x_I).



Relative Resistances

- The slope $-k_xa/k_ya$ determines the relative resistances of
- the two phases to mass transfer.
  - The distance AE is the gas-phase driving force ($y - y_I$).
  - The distance AF is the liquid-phase driving force ($x_I - x$).
  - If the resistance in the gas phase is very low, $y_I \sim y$.
    - Then, the resistance resides entirely in the liquid phase.
    - This occurs in the absorption of a slightly soluble solute in the liquid phase(a solute with a high K-value, e.g., CO_2).
    - Referred to as a liquid-film controlling process.
  - If the resistance in the liquid phase is very low, $x_I \sim x$.
    - Then, the resistance resides entirely in the gas phase.
    - This occurs in the absorption of a very soluble solute in the liquid phase (a solute with a low K-value, e.g., NH_3) .
    - Referred to as a gas-film controlling process.
- Important to know which of the two resistances is controlling
  - so that its rate of mass transfer can be increased by promoting turbulence and/or increasing the dispersion of the controlling phase.

Interfacial Composition Limitations: Overall MT Coefficients

- The composition at the interface between two phases is difficult to measure, so overall volumetric mass-transfer coefficients are defined in terms of overall driving forces between the two phases. Using mole fractions,

$$r = K_y a\left(y - y^*\right) = K_x a\left(x^* - x\right)$$

Overall mass-transfer coefficient for the gas phase

Overall mass-transfer coefficient for the liquid phase

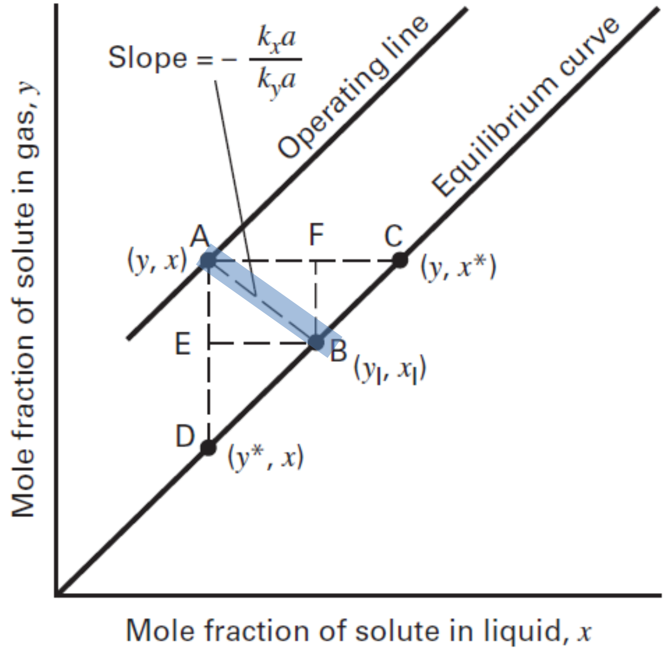
- y^* is the fictitious vapor mole fraction in equilibrium with the bulk liquid mole fraction, x, and x^* is the fictitious liquid mole fraction in equilibrium with the bulk vapor mole fraction, y.

Overall and Local MT Coefficients

- Overall MT coefficients can be expressed in terms of local phase coefficients:

$$\frac{1}{K_y a} = \frac{1}{k_y a} + \frac{1}{k_x a}\frac{(y_I - y^*)}{(x_I - x)}$$

$$\frac{1}{K_x a} = \frac{1}{k_x a} + \frac{1}{k_y a}\frac{(x^* - x_I)}{(y - y_I)}$$



- For dilute solutions:

$$\frac{(y_I - y^*)}{(x_I - x)} = \frac{\mathrm{ED}}{BE} = K$$

$$\frac{(x^* - x_I)}{(y - y_I)} = \frac{\mathrm{CF}}{FB} = \frac{1}{K}$$

⇒

$$\frac{1}{K_y a} = \frac{1}{k_y a} + \frac{K}{k_x a}$$

$$\frac{1}{K_x a} = \frac{1}{k_x a} + \frac{1}{K k_y a}$$

Packed Column Height

- Assume a dilute system in a countercurrent-flow absorption column.
- using a differential material balance for a solute being absorbed in a differential height of packing *dl* gives:

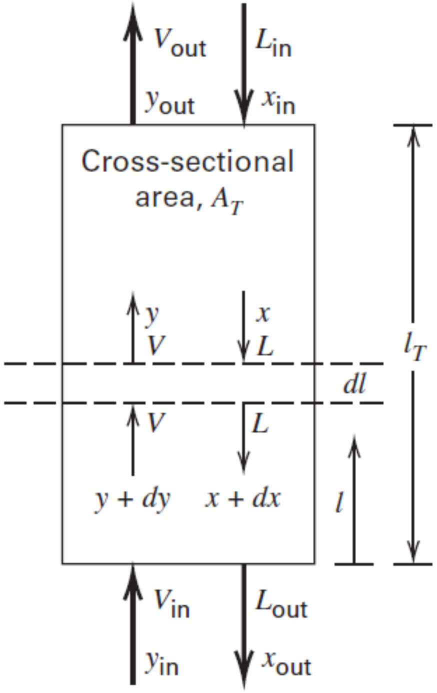
$$-V\,dy = K_y a\left(y - y^*\right) A_T\, dl$$

Cross sectional area of tower

- Assuming constant overall mass transfer coefficients and

$$\frac{K_y a A_T}{V}\int_0^{l_T} dl = \frac{K_y a A_T\, l_T}{V} = \int_{y_{out}}^{y_{in}} \frac{dy}{\left(y - y^*\right)}$$

$$l_T = \frac{V}{K_y a A_T}\int_{y_{out}}^{y_{in}} \frac{dy}{\left(y - y^*\right)}$$



Height and Number of Transfer Units (HTU and NTU)

- Chilton and Colburn suggested that:

$$l_T = \boxed{\frac{V}{K_y a A_T}} \boxed{\int_{y_{out}}^{y_{in}} \frac{dy}{(y - y^*)}} = H_{OG} N_{OG}$$

Overall number of (gas) transfer units (NTU)

Overall height of a (gas) transfer unit (HTU)

- Experimental data show that:
  - HTU varies less with *V* than does $K_y a$.
  - The smaller the HTU, the more efficient the contacting.
  - NTU represents the overall change in solute mole fraction divided by the average mole-fraction driving force.
  - The larger the NTU, the greater the time or area of contact required.

Packed Height from Linear Equilibrium Case

- The integral N_{OG} was integrated by Colburn, who used a linear equilibrium, $y^* = Kx$, to eliminate y^* and a linear solute material-balance operating line to eliminate x, yielding:

$$N_{OG} = \frac{\ln\left\{\left[(A-1)/A\right]\left[(y_{in} - Kx_{in})/(y_{out} - Kx_{in})\right] + 1/A\right\}}{(A-1)/A}$$

- Warning: a very sensitive calculation when A<0.9.

HTU and NTU vs. HETP and N_t

- NTU (N_{OG}) and HTU (H_{OG}) are not equal to the number of equilibrium stages, N_t, and HETP, respectively, unless the operating and equilibrium lines are straight and parallel.
- When the operating and equilibrium lines are straight but not parallel,

$$HETP = H_{OG}\frac{\ln(1/A)}{(1-A)/A}$$

$$N_{OG} = N_t\frac{\ln(1/A)}{(1-A)/A}$$

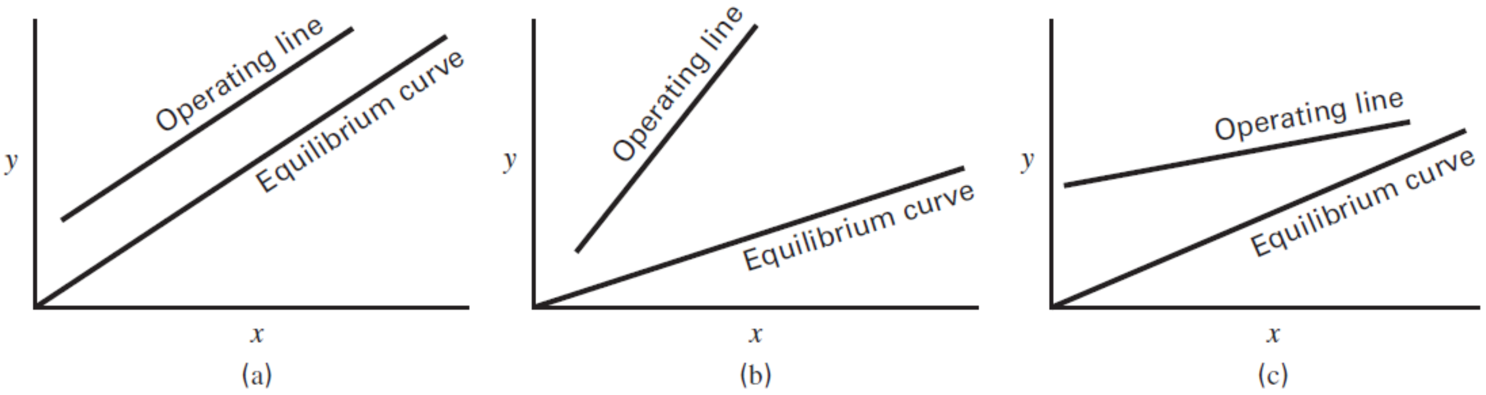
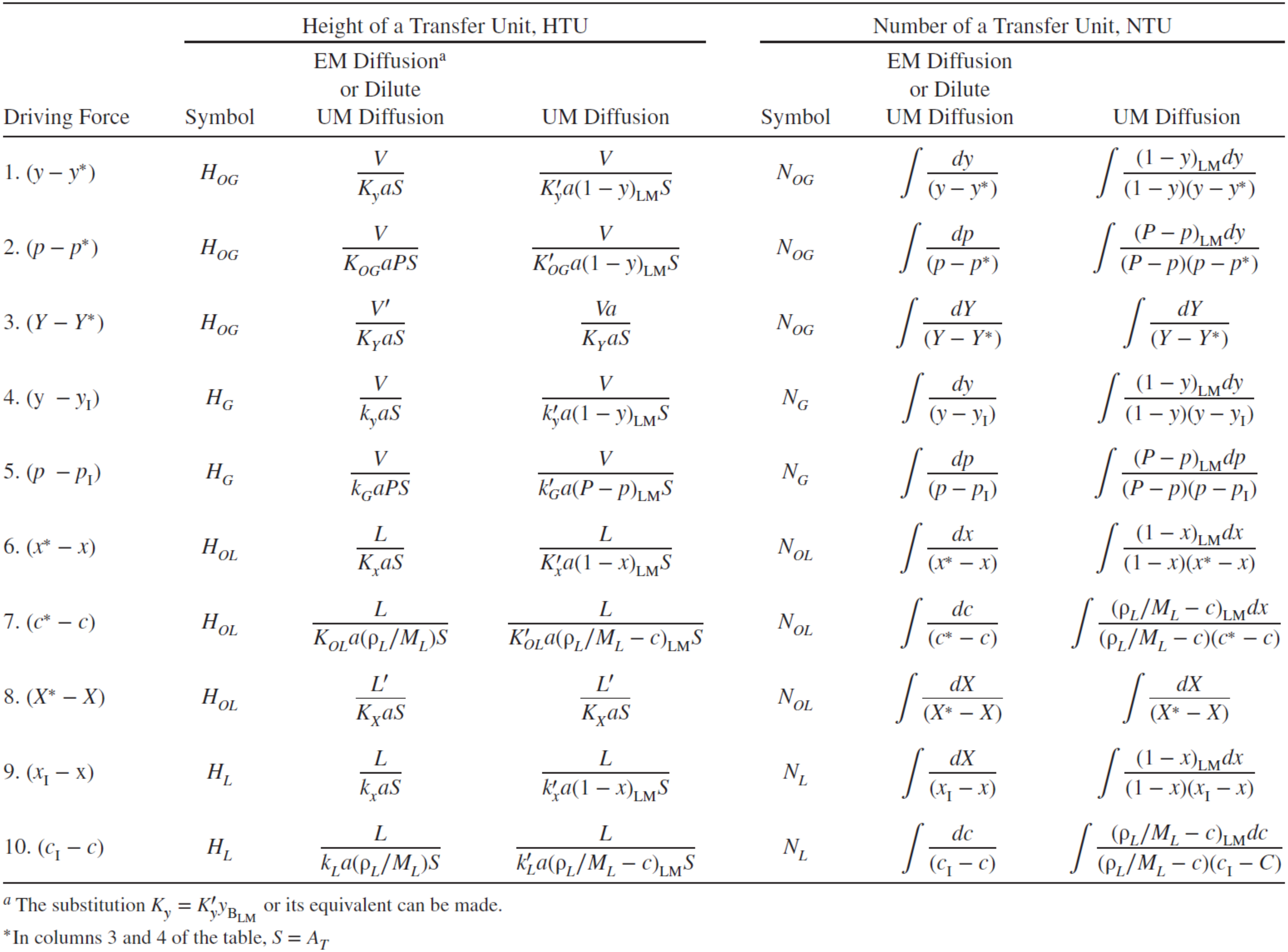


Figure 6.33 Relationship between NTU and the number of theoretical stages N_t: (a) NTU = N_t; (b) NTU > N_t; (c) NTU < N_t.

NTU and HTU: Alternative Mass-Transfer Coefficient Groupings



	Height of a Transfer Unit, HTU			Number of a Transfer Unit, NTU		
Driving Force	Symbol	EM Diffusion[a] or Dilute UM Diffusion	UM Diffusion	Symbol	EM Diffusion or Dilute UM Diffusion	UM Diffusion
1. $(y - y^*)$	H_{OG}	$\frac{V}{K_y aS}$	$\frac{V}{K'_y a(1-y)_{\mathrm{LM}} S}$	N_{OG}	$\int \frac{dy}{(y - y^*)}$	$\int \frac{(1-y)_{\mathrm{LM}} dy}{(1-y)(y-y^*)}$
2. $(p - p^*)$	H_{OG}	$\frac{V}{K_{OG} aPS}$	$\frac{V}{K'_{OG} a(1-y)_{\mathrm{LM}} S}$	N_{OG}	$\int \frac{dp}{(p - p^*)}$	$\int \frac{(P-p)_{\mathrm{LM}} dy}{(P-p)(p-p^*)}$
3. $(Y - Y^*)$	H_{OG}	$\frac{V'}{K_Y aS}$	$\frac{Va}{K_Y aS}$	N_{OG}	$\int \frac{dY}{(Y - Y^*)}$	$\int \frac{dY}{(Y - Y^*)}$
4. $(y - y_{\mathrm{I}})$	H_G	$\frac{V}{k_y aS}$	$\frac{V}{k'_y a(1-y)_{\mathrm{LM}} S}$	N_G	$\int \frac{dy}{(y - y_{\mathrm{I}})}$	$\int \frac{(1-y)_{\mathrm{LM}} dy}{(1-y)(y-y_{\mathrm{I}})}$
5. $(p - p_{\mathrm{I}})$	H_G	$\frac{V}{k_G aPS}$	$\frac{V}{k'_G a(P-p)_{\mathrm{LM}} S}$	N_G	$\int \frac{dp}{(p - p_{\mathrm{I}})}$	$\int \frac{(P-p)_{\mathrm{LM}} dp}{(P-p)(p-p_{\mathrm{I}})}$
6. $(x^* - x)$	H_{OL}	$\frac{L}{K_x aS}$	$\frac{L}{K'_x a(1-x)_{\mathrm{LM}} S}$	N_{OL}	$\int \frac{dx}{(x^* - x)}$	$\int \frac{(1-x)_{\mathrm{LM}} dx}{(1-x)(x^*-x)}$
7. $(c^* - c)$	H_{OL}	$\frac{L}{K_{OL} a(\rho_L/M_L)S}$	$\frac{L}{K'_{OL} a(\rho_L/M_L - c)_{\mathrm{LM}} S}$	N_{OL}	$\int \frac{dc}{(c^* - c)}$	$\int \frac{(\rho_L/M_L - c)_{\mathrm{LM}} dx}{(\rho_L/M_L - c)(c^*-c)}$
8. $(X^* - X)$	H_{OL}	$\frac{L'}{K_X aS}$	$\frac{L'}{K_X aS}$	N_{OL}	$\int \frac{dX}{(X^* - X)}$	$\int \frac{dX}{(X^* - X)}$
9. $(x_{\mathrm{I}} - x)$	H_L	$\frac{L}{k_x aS}$	$\frac{L}{k'_x a(1-x)_{\mathrm{LM}} S}$	N_L	$\int \frac{dX}{(x_{\mathrm{I}} - x)}$	$\int \frac{(1-x)_{\mathrm{LM}} dx}{(1-x)(x_{\mathrm{I}}-x)}$
10. $(c_{\mathrm{I}} - c)$	H_L	$\frac{L}{k_L a(\rho_L/M_L)S}$	$\frac{L}{k'_L a(\rho_L/M_L - c)_{\mathrm{LM}} S}$	N_L	$\int \frac{dc}{(c_{\mathrm{I}} - c)}$	$\int \frac{(\rho_L/M_L - c)_{\mathrm{LM}} dc}{(\rho_L/M_L - c)(c_{\mathrm{I}}-C)}$

[a] The substitution $K_y = K'_y y_{B_{\mathrm{LM}}}$ or its equivalent can be made.

*In columns 3 and 4 of the table, $S = A_T$

EXAMPLE 6.11 Height of an Ethanol Absorber.

Repeat Example 6.1 for a tower packed with 1.5-inch metal Pall rings. If $H_{OG} = 2.0$ ft, compute the required packed height.

Solution

From Example 6.1, $V = 180$ kmol/h, $L = 151.5$ kmol/h, $y_{\text{in}} = 0.020$, $x_{\text{in}} = 0.0$, and $K = 0.57$. For 97% recovery of ethyl alcohol by material balance,

$$y_{\text{out}} = \frac{(0.03)(0.02)(180)}{180 - (0.97)(0.02)(180)} = 0.000612$$

$$\mathcal{A} = \frac{L}{KV} = \frac{151.5}{(0.57)(180)} = 1.477$$

$$\frac{y_{\text{in}}}{y_{\text{out}}} = \frac{0.020}{0.000612} = 32.68$$

From (6-88),

$$N_{OG} = \frac{\ln\{[(1.477-1)/1.477](32.68) + (1/1.477)\}}{(1.477-1)/1.477}$$

$$= 7.5 \text{ transfer units}$$

The packed height, from (6-84), is $l_T = 2.0(7.5) = 15$ ft. N_t was determined in Example 6.1 to be 6.1. The 7.5 for N_{OG} is greater than N_t because the operating-line slope, L/V, is greater than the slope of the equilibrium line, K, so that Figure 6.33b applies.

EXAMPLE 6.12 Absorption of SO_2 in a Packed Column.

Air containing 1.6% SO_2 by volume is scrubbed with pure water in a packed column that is 1.5 m^2 in cross-sectional area and 3.5 m in packed height. Entering gas and liquid flow rates are 0.062 and 2.2 kmol/s, respectively. If the outlet mole fraction of SO_2 in the gas is 0.004 and the column temperature is near-ambient, with $K_{SO_2} = 40$, calculate (a) N_{OG} for the absorption of SO_2, (b) H_{OG} in meters, and (c) $K_y a$ for SO_2 in kmol/m^3-s-(Δy).

Solution

(a) The operating line is straight because the system is dilute in SO_2.

$$\mathcal{A} = \frac{L}{KV} = \frac{2.2}{(40)(0.062)} = 0.89$$

$$y_{\text{in}} = 0.016, \quad y_{\text{out}} = 0.004, \quad x_{\text{in}} = 0.0$$

From (6-88), we know that

$$N_{OG} = \frac{\ln\{[(0.89 - 1)/0.89](0.016/0.004) + (1/0.89)\}}{(0.89 - 1)/0.89}$$

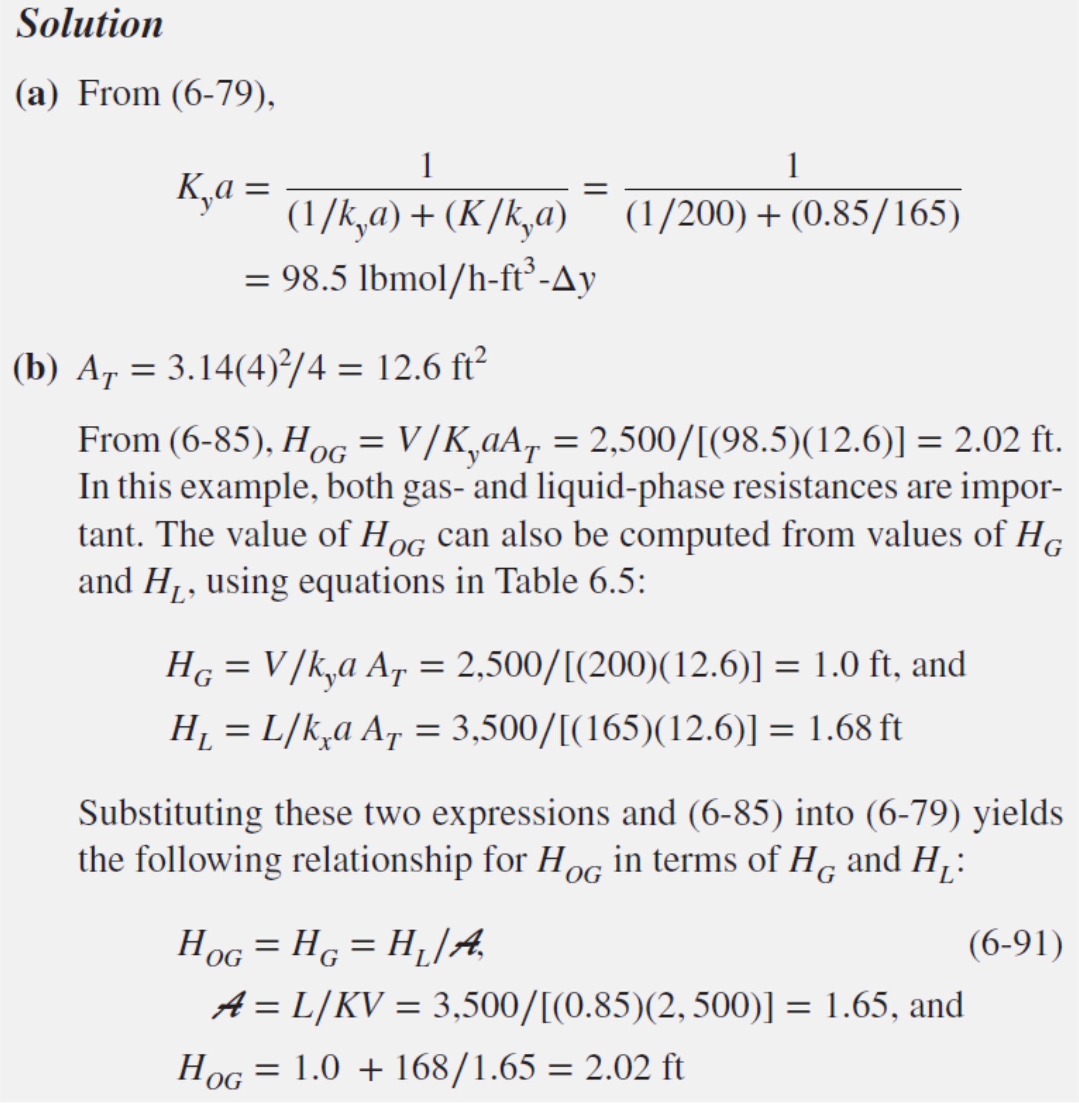
$$= 3.75$$

(b) $l_T = 3.5$ m. From (6-84), $H_{OG} = l_T/N_{OG} = 3.5/3.75 = 0.93$ m.

(c) $V = 0.062$ kmol/s, $A_T = 1.5\ m^2$. From (6-85), $K_y a = V/H_{OG}A_T$ $= 0.062/[(0.93)(1.5)] = 0.044$ kmol/m^3-s-(Δy).

EXAMPLE 6.13 Absorption of Ethylene Oxide.

A gaseous reactor effluent of 2 mol% ethylene oxide in an inert gas is scrubbed with water at 30°C and 20 atm. The gas feed rate is 2,500 lbmol/h, and the entering water rate is 3,500 lbmol/h. The column diameter is 4 ft, and the column is packed in two 12-ft-high sections with 1.5-inch metal Pall rings. A liquid redistributor is located between the packed sections. At column conditions, the *K*-value for ethylene oxide is 0.85, and the estimated values of $k_y a$ and $k_x a$ are 200 lbmol/h-ft³-Δy and 165 lbmol/h-ft³-Δx. Calculate: (a) $K_y a$ and (b) H_{OG}.



Solution

(a) From (6-79),

$$K_y a = \frac{1}{(1/k_y a) + (K/k_y a)} = \frac{1}{(1/200) + (0.85/165)}$$
$$= 98.5 \text{ lbmol/h-ft}^3\text{-}\Delta y$$

(b) $A_T = 3.14(4)^2/4 = 12.6 \text{ ft}^2$

From (6-85), $H_{OG} = V/K_y a A_T = 2{,}500/[(98.5)(12.6)] = 2.02$ ft. In this example, both gas- and liquid-phase resistances are important. The value of H_{OG} can also be computed from values of H_G and H_L, using equations in Table 6.5:

$$H_G = V/k_y a\, A_T = 2{,}500/[(200)(12.6)] = 1.0 \text{ ft, and}$$
$$H_L = L/k_x a\, A_T = 3{,}500/[(165)(12.6)] = 1.68 \text{ ft}$$

Substituting these two expressions and (6-85) into (6-79) yields the following relationship for H_{OG} in terms of H_G and H_L:

$$H_{OG} = H_G = H_L/\mathcal{A}, \qquad (6\text{-}91)$$
$$\mathcal{A} = L/KV = 3{,}500/[(0.85)(2,500)] = 1.65, \text{ and}$$
$$H_{OG} = 1.0\ + 168/1.65 = 2.02 \text{ ft}$$