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To Dot, Chuck, and Jennie

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Preface

The increasing challenges of climate change and a pandemic have brought a sense of urgency to finding solutions, many of which have a separations component. The continued development of industries such as biotechnology, nanotechnology, solar and wind power; the recognition that plastics need to be controlled; the increased importance of removing traces of compounds; the impressively rapid development of vaccines; and the need to reduce carbon dioxide emissions have brought new separations to the fore. Chemical engineers must understand and design new separation processes such as membrane separations, adsorption, and chromatography in addition to designing distillation, absorption, and extraction with more attention to reducing energy use.

The title of this book has evolved through its five editions. It began as *Equilibrium-Staged Separations*, but because membrane separations, adsorption, chromatography, and ion exchange were added, I changed the title of the second edition to *Separation Process Engineering* to reflect this broader coverage. The new title has been retained for the third, fourth, and fifth editions that featured a subtitle, *Includes Mass Transfer Analysis*, to reflect the new content in Chapter 15. The second edition was unavoidably longer than the first, the third edition was longer than the second, the fourth longer than the third, and, of course, the fifth edition is still longer. The most significant addition to the fifth edition is Chapter 18, “Melt Crystallization,” an important topic that has been essentially ignored in the United States. In addition to solid-liquid phase equilibrium, this chapter covers suspension, static layer, and falling-film layer melt crystallization and includes 33 questions and problems—27 of them new.

Additional extensive changes include the following:

- Chapter 2: New binary VLE equations and revision of material on simultaneous solutions
- Chapters 3 and 10: New sections on safety and fire hazards plus comments on dangerous chemicals throughout the problem sections
- Chapter 7: A new, more accurate correlation for the Gilliland correlation
- Chapter 8: New material on steam distillation
- Chapter 9: Three new examples, including new material on simple multicomponent batch distillation and residue curve analysis
- Chapter 10: Significant increase in coverage of tray efficiencies and packed column design
- Chapter 11: New equipment costs and increased discussion of energy reduction methods for distillation
- Chapter 13: Discussion of two additional hybrid processes with distillation plus application of the Kremser equation to fractional extraction with an example
- Chapter 14: A revised example of continuous washing, a new batch washing section, and coverage of the Shanks system for percolation leaching
- Chapter 15: An example of sublimation
- Chapter 17: New sections on deicing with eutectic systems, eutectic freeze concentration, and scaleup
- Chapter 19: New sections on forward osmosis and microfiltration

In addition, the previous single chapter on adsorption and ion exchange now comprises two chapters—20 and 21—with most of the new material in Chapter 21. Hybrid separations are extensively covered in Chapters 8, 12, 13, 17, 18, 19, and 20.

The more than 300 new questions and problems help students develop their problem-solving skills and discover new applications of theories and separation methods. Many of the problems were created and solved as I continued to teach this material, so a Solutions Manual is available to professors who adopt this textbook for their courses. A number of spreadsheet problems have been added, and the answers are provided in the Solutions Manual.

Because process simulators are used extensively in commercial practice, I have continued to include process simulation labs and homework problems in the chapter appendices. I taught the required three-credit, junior-level separations course at Purdue University as two lectures and a 2-hour computer lab every week. The computer lab included a lab test to assess the ability of the students to use the simulator. Although I use Aspen Plus as the simulator, any process simulator can be used. All the Aspen Plus labs have been updated to Version 12.1. The appendices to Chapters 2, 4, 5, 8, 9, 15, 17, and 19 have Excel spreadsheets, some of which use Visual Basic programs. I chose to use spreadsheets instead of a higher-level mathematical program because spreadsheets are universally available. The appendix to Chapter 21 includes tested instructions for eight labs delineating the operation of the commercial Aspen Chromatography simulator and all updated to Version 12.1. This new version is significantly more stable than version 8.8 used in the fourth edition.

The material in the fifth edition has been extensively tested in the required junior-level courses on separations and in a graduate-level elective at Purdue. Chapters 1 to 14 can be taught to sophomores, and all the material is suitable for seniors. The book is too long to cover in one semester, but more complete coverage is feasible in two quarters. If mass transfer is included, this text could easily be used for a two-semester sequence. If a single, three-credit semester course is available for separations, topics must be selected. One option is to use most of Chapters 1 to 14 and selected parts of Chapters 17, 19, or 20 for the required course and Chapters 15 to 21, including the Aspen Chromatography labs, for a senior-graduate-level elective course. Several course outlines are included in the Solutions Manual. Instructors may register at www.pearsonhighered.com for access to this book's Solutions Manual.

Register your copy of *Separation Process Engineering, Fifth Edition*, on the InformIT site for convenient access to updates and/or corrections as they become available. To start the registration process, go to informit.com/register and log in or create an account. Enter the product ISBN (9780137468041) and click Submit. Look on the Registered Products tab for an Access Bonus Content link next to this product, and follow that link to access any available bonus materials. If you would like to be notified of exclusive offers on new editions and updates, please check the box to receive email from us.

Acknowledgments

Many people were very helpful in the writing of the first edition. Dr. Marjan Bace and Prof. Joe Calo got me started writing. A. P. V. Inc., Glitsch Inc., and The Norton Co. kindly provided photographs. Chris Roesel and Barb Naugle-Hildebrand did the original artwork. The secretarial assistance of Carolyn Blue, Debra Bowman, Jan Gray, and Becky Weston was essential for completion of the first edition. My teaching assistants Magdiel Agosto, Chris Buehler, Margret Shay, Sung-Sup Suh, and Narasimhan Sundaram were very helpful in finding errors. Professors Ron Andres, James Caruthers, Karl T. Chuang, Alden Emery, and David P. Kessler, and Mr. Charles Gillard were very helpful in reviewing portions of the text. I also owe a debt to the professors who taught me this material: Lowell Koppel, who started my interest in separations as an undergraduate; William R. Schowalter, who broadened my horizons beyond equilibrium-staged separations in graduate school; and C. Judson King, who kept my interests alive while I was a professor and administrator through his articles, book, and personal example.

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For the fifth edition, I again thank a great group of reviewers: Priscilla Hill, Kent Knaebel, Stewart Slater, and Joe Shaeiwitz, who helped improve the book. I thank my editor, Malobika Chakraborty, for her assistance and encouragement to complete the book in a timely fashion.

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Nomenclature

Chapters 1 through 16

Note: Chapter 13 also has specialized nomenclature; see Table 13-1

a	interfacial area per volume, ft ² /ft ³ or m ² /m ³
a	constant in Eq. (2-11) and Table 2-4
a_j^l	interfacial area for heat transfer on stage j, m ²
$a_{\text{flow}}, a_{\text{heat}}, a_{\text{mass}}$	eddy diffusion parameters, Eqs. (15-61)
a_p	surface area/volume, m ² /m ³
$a_{p1}, a_{p2}, a_{p3}, a_{T1}, a_{T2}, a_{T6}$	constants in Eq. (2-28) and Table 2-5
A	LLE, solute, material being extracted
A	area, m ²
A, B	constants in Eq. (10-8b) to calculate pure component viscosities
A, B, C	constants in Antoine Eq. (2-9)
A, B, C, D	constants in matrix form of mass balances, Eqs. (6-6), (6-9), (6-12), (6-13), (12-46), and (12-48)
A, B, C, D, E	constants in Eq. (2-59a)
A_E, B_E, C_E, D_E	constants in matrix form of energy balances, Eqs. (6-28), (6-29a, b, c), (12-55a, b, c), (12-57), and (12-58)
A_{active}	active area of tray, ft ² or m ²
A_c	cross-sectional area of column, ft ² or m ²
A_d	downcomer area, ft ² or m ²
A_{du}	flow area under downcomer apron, Eq. (10-28), ft ²
A_f	area for flow, m ²
A_{hole}	area of holes in sieve plate, ft ²
A_I	interfacial area between two phases, ft ² or m ²
A_{mixer}	cross-sectional area of mixer, m ²
A_{mt}	area for mass transfer, m ²
A_{net}	net area available for vapor flow, Eq. (10-13), ft ² or m ²
A_{total}	total area in horizontal drum, ft ² or m ²
A_{valve}	area of valve in valve trays, ft ² , Eq. (10-35)
A_{vap}	area for vapor flow in horizontal drum, ft ² or m ²
Area	$= \int_{x_{\text{fm}}}^{x_{\text{F}}} \frac{dx}{y-x}$) batch distillation, Figure 9-4
b	constant in Eq. (2-11) and Table 2-4
b	equilibrium constant for linear equilibrium, $y = mx + b$, Eq. (12-9)
$b_{\text{flow}}, b_{\text{heat}}, b_{\text{mass}}$	eddy diffusion parameters, Eqs. (15-61) and (15-62)

B	bottoms flow rate, kmol/h or lbmol/h
B	solute
C	number of components
C	flow rate of direct cooling, kmol/h, lbmol/s, etc. Figure in Problem 3.D3
C	condensed material flow rate caused by subcooled inlet to column, kmol/h, lbmol/s, etc. Figure in Example 4-5
C	valve coefficient in valve trays
C_{BM}	bare module cost, Eqs. (11-8) to (11-10)
C_C	concentration of solute in continuous phase, kmol/m ³ continuous phase
C_C^*	concentration of solute in continuous phase in equilibrium with C_D , kmol/m ³
C_D	concentration of solute in dispersed phase, kmol/m ³ dispersed phase
C_{fL}	vapor load coefficient, Eq. (16-38) and Figure 16-7
C_A, C_B, C_m	molar concentrations, of A, B, and C_m = total concentration of mixture, kmol/m ³
C_o	orifice coefficient, Eqs. (10-24) and (10-25)
C_p	heat capacity, Btu/(lb·°F), Btu/(lbmol·°F), J/(g·°C), J/(mol·°C), etc.
C_p^o	base purchase cost as variable in Eq. (11-4) using values in Table 11-2
$C_{p,W}$	water heat capacity
$C_{p,size}$	packing size factor, Table 10-5
C_s	capacity factor at flood, Eq. (10-48)
$C_{sb,f}, C_{sb,flood}$	capacity factor at flood, Eqs. (10-9a) and (10-10) and Figure 10-15
$d_{hydraulic}$	hydraulic diameter for continuous phase flow, Eq. (13-58), m
d_i	impeller diameter, m
d_o	diameter sieve tray holes, in.
d_p, d_d	drop diameter, m
d_p^o	characteristic drop diameter, Eq. (16-97b), m
d_{pack}	packing diameter, m or ft
$d_{settler}$	diameter of horizontal settler, m
d_{tube}	tube diameter, m
df	damping factor, $df \leq 1$
D	diffusivity, Fickian, m ² /s or ft ² /h
D^o	infinite dilution Fickian diffusivity, m ² /s
\bar{D}	diffusivity, Maxwell-Stefan, m ² /s or ft ² /h
\bar{D}^o	infinite dilution Maxwell-Stefan diffusivity, m ² /s or ft ² /h
D	distillate flow rate, kmol/h or kg/h
D	diluent, chemical solute is dissolved in the feed
D	amount of distillate, kmol (Chapter 9)
D	diameter of sphere, m (Chapter 15)
D, Dia	diameter of column, ft or m
D'_{col}	column diameter, see Table 16-2, ft
D_{rate}	distillate flow rate in batch distillation, kmol/h, Eq. (9-30a)
D_{eddy}	eddy diffusivity, Eqs. (16-111a) and (16-111b), m/s
D_{min}	minimum diameter horizontal flash drum, Eq. (2-64b)
D_{rate}	distillate flow rate, kmol/h (Chapter 9)
D_{total}	total amount of distillate (Chapter 9), kg, lb, kmol, lbmol, etc.
$D_{V,i}$	vapor diffusivity, Section 16.8, cm ² /s, m ² /s
e	absolute entrainment, mol/h

E	vaporization efficiency for batch steam distillation, Eq. (9-19), dimensionless
E	extract flow rate (Chapters 13 and 14), kg/h
\bar{E}	extract flow rate in bottom half of extractor with center feed, kg/h
\hat{E}	mass extract in batch extraction, kg
E_j^V	energy transfer rate on stage j from bulk liquid to bulk vapor, J/s, Eq. (16-106b)
E_k	value of energy function for trial k, Eq. (2-48)
E_{MD}	Murphree dispersed phase efficiency in extraction, Eqs. (13-51) and (16-84)
E_{ML}, E_{MV}	Murphree liquid and vapor efficiencies, Eqs. (4-58) and (4-59)
E_0	activation energy, Kcal/mol
E_o	overall efficiency, Eqs. (4-57) and (10-1)
E_{pt}	point efficiency, Eq. (10-4c) or Eqs. (16-76a)
\hat{E}_t	holdup extract phase in tank plus settler, kg
(EB_j)	energy balance function, essentially Eq. (6-28)
f	friction factor
f_{AB}	friction coefficient between molecules A and B
f_c	fraction condensed per mol subcooled reflux, Eq. (4-62)
$f = V/F$	fraction vaporized or fraction of feed that is vapor
f	fractional approach to flooding
$f(x)$	equilibrium function for y (Chapter 9)
$f_k(V/F)$	Rachford-Rice function for trial K, Eqs. (2-41) to (2-43)
F	packing factor, Tables 10-3 and 10-4, Eq. (10-39)
F	degrees of freedom, Eq. (2-4)
F	charge to still pot (Chapter 9), kg, lb, kmol, lbmol, etc.
\hat{F}	mass of feed in batch extraction, kg
F	feed flow rate, kmol/h, lbmol/h, kg/h, etc.
F_D	diluent flow rate (Chapter 13), kg/h
F_{lv}, FP	$\frac{W_L}{W_V} \sqrt{\frac{\rho_L}{\rho_V}} = \frac{L'}{G'} \sqrt{\frac{\rho_L}{\rho_V}}$, flow parameter
F_m	material factor for cost, Table 11-5
F_p , pressure vessels	pressure factor for cost, Eq. (11-7) and Table 11-4
F_p , reboilers and condensers	pressure factor, Eq. (11-4) and Table 11-2
$[F_{old}]$	combined matrix of all functions, Eq. (6-40)
F_q	quantity factor for cost, Eq. (11-4) and Table 11-2
F_S, F_{solv}	flow rate solvent (Chapters 13 and 14), kg solvent/h
F_{solid}	solids flow rate in leaching (Chapter 14), kg insoluble solid/h
F_{weir}	weir modification factor, Eq. (10-26) and Figure 10-21
$FR_{A,dist}$	fractional recovery of A in distillate, Eqs. (3-5)
$FR_{B,bot}$	fractional recovery of B in bottoms, Eqs. (3-5)
gap	gap from downcomer apron to tray, Eq. (10-28), ft
g	acceleration due to gravity, 32.2 ft/s ² , 9.81 m/s ²
g_c	conversion factor in English units, 32.2 ft · lbf/(lbf · s ²)
G	flow rate insoluble carrier gas, kmol/h or kg/h, Eq. (12-41a)
G'	gas flux, lb/(s ft ²) for Eq. (10-40b)
h	pressure drop in head of clear liquid, in. liquid
h	height of liquid on stage (Chapter 16), ft
h	total height of packing in a section, m or ft, Eq. (16-4)
h	height of liquid in mixer, m
h	length of horizontal flash drum, m or ft, Eq. (2-65)
h	liquid enthalpy, kcal/kg, kJ/kmol, Btu/lbmol, etc.

h	increment size in Euler's method = Δt , Eqs. (9-A2) and (9-A3)
h	interval in Gaussian quadrature formula, Eq. (9-12b)
\tilde{h}	pure component enthalpy, kcal/kg, kJ/kmol, Btu/lbmol, etc.
h_0	reflux enthalpy, kcal/kg, kJ/kmol, Btu/lbmol, etc.
h_B	enthalpy of bottoms product, kcal/kg, kJ/kmol, Btu/lbmol, etc.
h_{cl}	calculated height of clear liquid on tray, Eq. (16-107c)
h_{crest}	height of liquid crest over weir, in., Eq. (10-26)
h_D	enthalpy of distillate product, kcal/kg, kJ/kmol, Btu/lbmol, etc.
h_{dc}	head of clear liquid in downcomer, Eq. (10-23), in.
h_{du}	frictional loss for flow liquid under downcomer, in. clear liquid, Eq. (10-27)
h_f	enthalpy of liquid leaving feed stage, kcal/kg, kJ/kmol, Btu/lbmol, etc.
h_F	feed enthalpy (liquid, vapor, or two phase), kcal/kg, kJ/kmol, Btu/lbmol, etc.
$h_{heat\ transfer}$	heat transfer coefficient, kW/(K·m ²), BTU/(h·ft ² ·°F)
h_L	clear liquid height on stage, m or cm
h_L, H	molar holdup of liquid on tray
h_o	hole diameter for sieve plates, in.
h_p	packing height, ft or m
h_{total}	height of flash drum, ft or m
h_{weir}	height of weir, in.
$h_{\Delta p, valve}$	head loss in inches clear liquid for valve trays, Eq. (10-34a)
H, H_B	volatility form of Henry's law constant, Eqs. (8-5), (8-6), (12-1), (12-3), (12-4) and (12-37)
H	height of staged column, ft, m, Eq. (11-17a)
H_{org}	Henry's law constant for organic in water phase
H_w	Henry's law constant for water in organic phase
H_{sol}	solubility form of Henry's law constant, Eqs. (12-5) and (12-37)
H	vapor enthalpy, kcal/kg, kJ/kmol, Btu/lbmol, etc.
$H_{v,i,j}$	partial molar enthalpy of component i in vapor on stage j, kJ/kmol, BTU/lbmol, etc.
H_{tank}	height of tank, m, ft
Head space	height of vapor space above liquid in tank, m, ft
HETP	height equivalent to a theoretical plate, ft or m, Eq. (10-36)
HETP _{practical}	HETP value including height taken by distribution equipment, Eq. (10-47a)
HETP _{tray}	equivalent HETP for trays, Eq. (10-47b)
$H_{t,OD}$	overall height of transfer unit, driving force in concentration units, Eq. (16-83a analog), ft or m
H_G	$= \frac{V}{k_y a A_c}$ height of gas-phase transfer unit, ft or m, Eq. (16-6a, b)
$H_{G,E}, H_{L,E}$	H_G or H_L in the enriching section
H_L	$= \frac{L}{k_x a A_c}$ height of liquid-phase transfer unit, ft or m, Eq. (16-11a)
H_{O-ED}	$= (Q_D/A_{mixer})(\rho_D/MW_D)/(K_{O-ED}a)$, LLE, height overall dispersed phase transfer unit, m, Eq. (16-83a)
H_{OG}	$= \frac{V}{K_y a A_c}$ height of overall gas-phase transfer unit, ft or m, Eq. (16-22a)

H_{OL}	$= \frac{L}{K_x a A_c}$ height of overall liquid-phase transfer unit, ft or m, Eq. (16-25a)
H_{O-raf}	$= (Q_{raf}/A_c)(\rho_{raf}/MW_{raf})/(K_{O-raf}a)$, height overall raffinate transfer unit, m, Eq. (16-82b)
HTU	height of a transfer unit, ft or m
j_D, j_H	Chilton-Colburn analogy j-functions, Eqs. (15-63b)
$J_{A,z}$	molar flux in direction with respect to molar average velocity of fluid, mole A/(s·m ²)
k_B	Boltzmann's constant, J/k (see Appendix C)
k_c	mass transfer coefficient with concentration driving force, m/s, Eq. (15-24b), Table 16-1
$k_{conduction}$	thermal conductivity, J/(m·s·K)
\bar{k}_x, \bar{k}_y	individual mass transfer coefficients in liquid and vapor phases, see Table 16-1
k'_y	mass transfer coefficient in concentrated solutions, Eq. (15-32f)
k_x, k_y	individual mass transfer coefficient in molar units
$k_{x,c}, k_{x,D}$	individual mass transfer coefficients in continuous and dispersed phases, kg/(s·m ³), lbm/(s·ft ³), kmol/(s·m ³), or lbmol/(s·ft ³)
k_{LD}, k_{LC}	individual mass transfer coefficients in continuous and dispersed phases with driving force in concentration units, m/s
k_L, k_V	individual liquid and vapor mass transfer coefficients, distillation, Eqs. (16-108) to (16-110), m/s
k	mass transfer coefficient in Maxwell-Stefan analysis, $\bar{D}/\Delta z$, m/s
K	$K = C_{sb,f} \left(\frac{\sigma}{20}\right)^{0.2}$ Eq. (10-9a)
K_d	y/x, distribution coefficient for dilute extraction, Eq. (13-1a) and Table 13-3
K, K_i	y_i/x_i , equilibrium vapor-liquid ratio
$K_{i,V-L1}, K_{i,V-L2}$	vapor K values for three-phase system, Eqs. (2-54a) and (2-54b)
$K_{i,L1-L2}$	liquid-liquid distribution coefficient for three-phase system, Eq. (2-54c)
K_{drum}	parameter to calculate u_{perm} for flash drums, Eq. (2-61)
$K_{horizontal}$	K_{drum} for a horizontal drum, Eq. (2-64d)
$K_{vertical}$	K_{drum} for a vertical drum, Eq. (2-62a)
$K_{v,closed}, K_{v,open}$	parameters for head loss in valve trays, Eqs. (10-34a) and (10-34b)
K_x, K_y	overall mass transfer coefficient, liquid and vapor, Eqs. (1-5), (15-29), and (15-31), lbmol/(ft ² ·h), or kmol/(h·m ²)
K_{LD}	overall mass transfer coefficient dispersed phase in concentration units, Eq. (16-80b analog), m/s
K_{O-ED}	overall dispersed phase LLE mass transfer coefficient, Eq. (16-80a), kg/(s·m ³), kmol/(s·m ³)
K_{O-raf}	overall raffinate LLE mass transfer coefficient, Eq. (16-82b), kmol/(s·m ³)
K_1, K_2, K_3	Factors from Table 11-2 for Eq. (11-4)
$\ell_{i,j}$	$\ell_{i,j} = x_{i,j} L_j$, liquid component flow rate component i, stage j, kmol/h, lbmol/h, etc., Eq. (6-3)
l_{weir}	weir length, ft, m
ℓ	lateral coordinate across tray, Figure 16-10, ft or m
L	length, ft or m
L	liquid flow rate, kmol/h or lbmol/h
L'	liquid flux, lb/(s·ft ²) for Eq. (10-40b)

L_g	liquid flow rate in gal/min (Chapter 10)
L_F	liquid flow rate of feed, kmol/h, lbmol/h, etc.
L_{\min}	minimum liquid flow rate in enriching section, kmol/h, lbmol/h, etc.
L_0/D	external reflux ratio, dimensionless
m	linear equilibrium constant, $y = mx + b$
m	local slope of equilibrium curve, Eqs. (10-4b) and (15-30b)
m_{CD}	LLE: slope of equilibrium curve in mass or mole fractions, Eq. (16-80c)
$m_{CD,conc_units}$	LLE: slope of equilibrium curve in concentration units, Eq. (16-80c analog)
M	flow rate of mixed stream (Chapters 8 and 13), kg/h
M	multiplier times $(L/D)_{\min}$ (Chapter 7)
M	ratio $HETP_{\text{practical}}/HETP_{\text{packing}}$, Eq. (10-47a); random: $M = 1.1$; structured: $M = 1.2$
MW	molecular weight
\overline{MW}	average molecular weight
n	exponent in Eq. (11-5) relating equipment cost to size.
n	moles
n	number of drops
n_G	$= \int_{y_{A,in}}^{y_{A,out}} \frac{dy_A}{y_{A,I} - y_A}$ number gas-phase transfer units, distillation and dilute absorption, Eq. (16-7)
n_G	number gas-phase transfer units for concentrated absorbers, Eq. (16-52)
n_L	$= \int_{x_{A,out}}^{x_{A,in}} \frac{dx_A}{x_A - x_{A,I}}$ number liquid-phase transfer units, distillation and dilute absorption, Eq. (16-11a)
n_{O-ED}, n_{O-EC}	number overall extraction transfer units in dispersed and continuous phases, Eq. (16-81a); (see also Eq. (16-83b))
n_{OG}	$= \int_{y_{A,in}}^{y_{A,out}} \frac{dy_A}{y_A^* - y_A}$ number overall gas-phase transfer units, distillation and dilute absorption, Eq. (16-23)
n_{OL}	$= \int_{x_{A,in}}^{x_{A,out}} \frac{dx_A}{x_A - x_A^*}$ number overall liquid-phase transfer units, distillation and dilute absorption, Eq. (16-26)
n_{org}	moles organic in vapor in steam distillation
n_w	moles water in vapor in steam distillation
N, ω	impeller revolutions per second
N	number of stages
N_A	molar flux of A, lbmol/(h·ft ²) or kmol/(h·m ²)
N_p, N_{feed}	feed stage
$N_{Li,j}$	transfer to liquid from vapor on stage j, mol component i/s
$N_{Vi,j}$	transfer to vapor from liquid on stage j, mol component i/s
N_{\min}	number of stages at total reflux
$N_{F,\min}$	estimated feed stage location at total reflux
N_{Po}	power number, Eq. (13-46)
N_{tOD}	LLE, number of overall transfer units, driving force in concentration units, Eq. (16-81a analog)

Nu	Nusselt number, Eq. (15-46g)
NTU	number of transfer units
O	total overflow rate in washing, kg/h
\hat{O}	amount of overflow liquid in batch washing, kg
p	pitch of sieve plate holes, m
P, P _{tot}	pressure, atm, kPa, psi, bar, etc.
P _c	critical pressure, atm, kPa, psi, bar, etc.
P _{drum}	drum pressure, atm, kPa, psi, bar, etc.
\bar{P} , P _B	partial pressure, atm, kPa, psi, bar, etc.
P	number of phases, Eq. (2-4)
P	power, W, kW
Pe	dimensionless Peclet number in terms of molecular diffusivity, Eq. (15-46c)
Pe	dimensionless Peclet number in terms of eddy diffusivity, Eq. (16-111a)
Per _f	flow perimeter, Figure 13-33B, m
Pr	$= \mu C_p / k_{\text{conductivity}}$, dimensionless Prandtl number, Eq. (15-46f)
q	$L_F / F = (L - L) / F$, feed quality
q	volumetric flow rate/plate width, m ² /s
Q	amount of energy transferred, Btu/h, kcal/h, J/s, W, etc.
Q _c	condenser heat load, Btu/h, kcal/h, J/s, etc.
Q _c , Q _C	volumetric flow rate continuous phase, m ³ /s
Q _{col}	heat loss from distillation column, Btu/h, kcal/h, J/s, etc. Adiabatic has Q _{col} = 0.
Q _d , Q _D	volumetric flow rate dispersed phase, m ³ /s
Q _{flash}	heat loss from flash drum, Btu/h, kcal/h, J/s, etc. Adiabatic has Q _{flash} = 0.
Q _I	intermediate reboiler heat load, Btu/h, kcal/h, J/s, etc.
Q _L	volumetric flow rate of liquid, m ³ /s
Q _R	reboiler heat load, Btu/h, kcal/h, J/s, W, etc.
Q _{Raf} /A _c	LLE = superficial velocity of raffinate in the column, m/s, ft/h, etc., Eq. (16-82b)
Q _z /A	heat flux in z direction, J/(m ² ·s)
r	radius of column, ft or m
r _{i,j}	reaction component i on stage j, Eq. (8-32)
R	gas constant, 1.9859 cal/(mol·K) or 8.314 m ³ Pa/(mol·K), Appendix C
R	raffinate flow rate (Chapters 13 and 14), kg/h
\bar{R}	raffinate flow rate in bottom section of extractor with center feed, kg/h
R	reagent in reactive absorption, Eqs (12-60a, b) and Figures 12-15 and 12-16B
R _A	solute radius, m
\hat{R}	mass raffinate in batch extraction, kg
\hat{R}_t	holdup raffinate phase in tank plus settler, kg
Re	dimensionless Reynolds number, Eq. (15-46b)
Re _{drop}	$= (\rho_c du_t / \mu_c)$ where u _t is the terminal velocity, Eq. (16-98c)
Re _{settler}	Reynolds number for settler, Eq. (13-56a)
S	solvent, in Chapters 12, 13, and 14 separating agent added for separation
S	steam flow rate into distillation column, kmol/s, kmol/h, lbmol/s, etc. Figure in Problem 3.D2.
S	flow rate nonvolatile solvent, kmol/h or lbmol/h

S/G	(kmol/h nonvolatile solvent)/(kmol/h insoluble carrier gas)
S	solvent flow rate, kg/h
S	kg dry solids/h, Eq. (14-5)
S	tray spacing, in., Eq. (10-47b)
S	moles second solvent in constant-level batch distillation
S	size factor from Table 11-2 for Eq. (11-4) or size factor from Table 11-3 for Eq. (11-5)
\hat{S}	mass of solvent in batch extraction, kg
S_{AB}	separating power of cascade, Eq. (13-18c)
S_L, S_V	flow rates of liquid and vapor sidestreams, kmol/h
Sc_L, Sc_V	Schmidt number for liquid or vapor = $\mu/(\rho D)$
Sh_c, Sh_x, Sh_y	dimensionless Sherwood numbers, Eqs. (15-46a) and (16-98)
St_c, St_x, St_y	dimensionless Stanton numbers, Eq. (15-46d)
t	time, s, min, or h
t_{batch}	period for batch distillation, Eq. (9-30a)
t_{down}	down time in batch distillation
$(t_{f,95} - t_0)$	residence time in extractor for 95.0% extraction, Eq. (16-104), s
t_L, t_V	average residence time per pass for liquid and vapor, s
$\bar{t}_{L,residence}$	liquid residence time, Eq. (16-111c), s
$t_{residence,dispersed}$	residence time of dispersed phase in settler, s
t_{op}	operating time in batch distillation
t_{res}	residence time in downcomer, Eq. (10-30), s, Table 10-2
t_{tray}	tray thickness, in.
T	temperature, °C, °F, K, or °R
T	time to fill drum with liquid, Eq. (2-69)
T_{BP}	bubble point temperature, °C, °F, K, or °R
T_{DP}	dew point temperature, °C, °F, K, or °R
T_{jIL}, T_{jIV}	liquid and vapor temperatures on stage j at the interface, °C, °F, K, or °R
T_{ref}	reference temperature, °C, °F, K, or °R
TAC	total annual cost, \$/y, Eq. (11-22)
u	vapor velocity, cm/s or ft/s
u_{flood}	flooding velocity, cm/s or ft/s, Eq. (10-8)
u_{op}	operating velocity, cm/s or ft/s, Eq. (10-11)
u_{perm}	permissible vapor velocity, cm/s or ft/s, Eq. (2-58)
$u_{t,hindered}$	hindered settling velocity, cm/s or ft/s, Eq. (13-54)
$u_t, u_{t,Stokes}$	Stokes law terminal velocity, cm/s or ft/s, Eq. (13-53)
U	underflow liquid rate, Eq. (14-5d), kg/h
\hat{U}	amount underflow liquid in batch washing, Eq. (14-12), kg
U	overall heat transfer coefficient, kW/(K·m ²), BTU/(h·ft ² ·°F), values Table 11-7
U_a	superficial vapor velocity in active area of tray, m/s
U_{nf}	= u_{flood} , Figure 10-15, ft/s or cm/s
U_1, U_2, \dots, U_N	Unknowns solved for by Thomas algorithm, Table 6-1
v	superficial vapor velocity, ft/s
$v_{c,flood}, v_{d,flood}$	continuous- and dispersed-phase flooding velocities, m/s
v_o	vapor velocity through holes, Eq. (10-19), ft/s
$v_{o,bal}$	velocity where valve in valve tray is balanced, Eq. (10-36)
v_A, v_B	component transfer velocities, m/s, Eqs. (15-15e) and (15-15f)
$v_{i,j}$	$v_{i,j} = y_{i,j} V_j$, vapor component flow rate, component i, stage j, kmol/h, lbmol/h, etc.
v_{ref}	reference or basis velocity, m/s, Eqs. (15-15c) and (15-15d)

$V_{\text{ref,mass}}, V_{\text{ref,mol}}, V_{\text{ref,vol}}$	reference velocities in different units, m/s, Eqs. (15-17c1) to (15-17c3)
v_y	vertical velocity, m/s
\bar{V}	vapor flow rate, kmol/s, kmol/h, lbmol/s, lbmol/h, etc.
\bar{V}	stripping section vapor flow rate, kmol/s, kmol/h, lbmol/s, lbmol/h, etc.
\bar{V}/\bar{B}	boilup ratio, dimensionless
V_i	molal volume, m ³ /mol, Table 13-4
V_A	molar volume solute at normal boiling point, m ³ /kmol
V_F	vapor flow rate of feed, kmol/h, lbmol/h, etc.
V_{liq}	volume of liquid in drum or tank, m ³
V_{max}	maximum vapor flow rate, kmol/h or lbmol/h
V_{min}	minimum vapor flow rate in enriching section, kmol/h, lbmol/h, etc.
\bar{V}_{min}	minimum vapor flow rate in stripping section, kmol/h, lbmol/h, etc.
V_{mixer}	volume of liquid in mixing tank, m ³
V_{settler}	volume settler, m ³
V_{tank}	volume tank, m ³
V_{surge}	surge volume in flash drum, Eq. (2-62), ft ³
(V1), (V2), (V3)	intermediate variables in Thomas algorithm, Table 6-1
VP	vapor pressure, same units as p
w	plate width, m
W	charge in still pot of batch distillation, kg, lb, kmol, lbmol, etc. (Chapter 9)
W_{final}	charge remaining in still pot at end of batch distillation, kg, lb, kmol, lbmol, etc., Chapter 9
W_L	liquid flow rate, kg/h or lb/h
W_L	liquid mass flux, lb/s ft ² or lb/h ft ² (Chapter 16)
W_V	vapor flow rate, kg/h or lb/h
W_{valve}	valve weight, lb
x	weight or mole fraction in liquid
x	LLE raffinate or feed weight fractions (see Table 13-1)
x	weight fraction in underflow in washing (Chapter 14)
x^*	equilibrium mole fraction in liquid
$x_{A,\text{az}}, x_{B,\text{az}}$	mole fractions binary azeotrope, Table 2-4 and Eqs. (2-10a) and (2-10b)
$x_{A\Delta}, x_{B\Delta}$	coordinates of difference or delta point, Eqs. (13-39a, b)
$x_{A,M}, x_{D,M}$	coordinates of mixed stream, Eqs. (13-31a, b) or (13-37a, b)
x_B, x_{bot}	mole fraction MVC in bottoms product
x_D, x_{dist}	mole fraction MVC in distillate
$x_{D,\text{avg}}$	average mole fraction MVC in distillate at end of batch
$x_{D1,\text{avg}}, x_{D2,\text{avg}}, \dots$	mole fractions of distillate collected
x_D	LLE: weight or mole fraction of solute A in the dispersed phase
x_D^*	fraction that would be in equilibrium with continuous-phase fraction A, Eq. (16-80a)
x_{fin}	final mole fraction MVC in still pot at end of batch
$x_{i,n}, x_{i,n+1}$	trials for integration, Eq. (9-A4)
x_I	intersection point of top and bottom operating lines, Eq. (4-38a)
x_I	interfacial mole fraction in liquid
x_N^*, x_{in}^*	liquid mole fraction in equilibrium with inlet gas, Eq. (12-35)
x_{mid}	mid-point in Gaussian quadrature formula, Eq. (9-12c)
x_0^*, x_{out}^*	liquid mole fraction in equilibrium with outlet gas, Eqs. (12-35) and (16-35b)

$X_{\text{org in org}}$	liquid mole fraction of organic in organic phase
$X_{\text{org in w}}$	liquid mole fraction of organic in water phase
$X_{\text{pot,org}}$	mole fraction organic in still pot in batch steam distillation
$X_{\text{w in org}}$	liquid mole fraction of water in organic phase
$X_{\text{w in w}}$	liquid mole fraction of water in water phase
X_W	liquid mole fraction of first solvent in still pot for constant-mole batch distillation, Eq. (9-16c)
X_s, X_{side}	mole fraction of MVC in sidestream
X	LLE weight ratio solute in immiscible diluent, Eq. (13-23)
X	mol solute in liquid/(mol pure solvent S)
X	$[L/D - (L/D)_{\text{min}}]/(L/D + 1.0)$ Gilliland's correlation, Eq. (7-34b)
X	kg solute/kg insoluble solid (Chapter 14)
y	weight or mole fraction in vapor
y	weight fraction in overflow in washing (Chapter 14)
y	vapor mole fraction of first solvent withdrawn in constant-mole batch distillation, Eq. (9-16c)
y	LLE extract or solvent stream weight fractions (see Table 13-1)
$y_{A,\text{ref}}, y_{B,\text{ref}}$	fractions to calculate velocity of center of total flux, Eq. (15-17d, e)
y_D	vapor mole fraction MVC in distillate
y_I	intersection point of top and bottom operating lines, Eq. (4-38b)
y_{vol}	volume fraction in vapor
y^*	equilibrium mole fraction in vapor
$y_{N+1}^*, y_{\text{in}}^*$	vapor mole fraction in equilibrium with outlet liquid in countercurrent system, Eq. (12-27)
$y_{1, \text{out}}^*, y_{\text{out}}^*$	vapor mole fraction in equilibrium with inlet liquid in countercurrent system, Eqs. (12-27) and (16-35a), or in equilibrium with outlet liquid in cocurrent contactor, Eq. (16-70b)
y_{lm}	log mean difference, Eqs. (15-32d) and (16-44)
y_I	interfacial mole fraction in vapor
y_S	mole fraction MVC in entering steam
y_{side}, y_s	mole fraction MVC in side stream
$\bar{y}_{j+1} = y_{j+1,\text{avg}}$	average vapor mole fraction leaving stage below
Y	LLE, weight ratio solute in immiscible solvent, Eq. (13-23)
Y	(moles solute in gas)/(moles noncondensable carrier gas)
Y	$= [N - N_{\text{min}}]/[N + 1]$, Gilliland's correlation, Eq. (7-34a)
Y	kg solute/kg solvent (Chapters 13 and 14)
z	weight or mole fraction in feed
z	axial distance in bed (Chapters 15 and 16)
z_1	distance from downcomer exit to weir, m
Z	(lb solution)/(lb oil-free solids) in underflow (Chapter 14)

Greek letters

α, β	parameters for pressure drop equation in randomly packed columns, Eq. (10-40b), Table 10-3
α_{AB}	Greek lowercase letter alpha, K_A/K_B , relative volatility
$\alpha_{AB,L}$	modified relative volatility, $x_A = 0$ to $x_{A,Az}$, Eq. (2-10a) and Table 2-4
$\alpha_{BA,R}$	modified relative volatility, $x_B = 0$ to $x_{B,Az}$, Eq. (2-10b) and Table 2-4
$\alpha_{BA,\text{Reb}}$	relative volatility calculated at reboiler
α_{AB}	LLE, separation factor, Eqs. (13-1b) and (13-1c)
α_{thermal}	thermal diffusivity = $k_{\text{conduction}}/(\rho C_p)$, m^2/s
β	Greek lowercase letter beta, $A_{\text{hole}}/A_{\text{active}}$

γ	Greek lowercase letter gamma, activity coefficient
γ	surface tension, dynes/cm
δ	Greek lowercase letter delta, thickness of mass transfer film or thickness of falling film, m
δ_p	characteristic dimension of packing, in., Eq. (10-39)
δ_i	solubility parameter, Eqs. (13-2) and Table 13-4
$\bar{\delta}$	weighted average value of δ_i , Eq. (13-2b)
Δ	Greek capital letter delta, change in variable or difference operator
Δ	difference flow rate, Eqs. (13-38a) or (13-42) and (14-15b)
Δ	difference point
Δ_{\min}	difference point corresponding to minimum S/F, Figure 13-29
ΔH	steady-state height of dispersion band in settler Figure 13-33B, cm, m
Δp	pressure drop inches water/ft packing, Eq. (10-40b)
ΔT_j	temperature changes for matrix energy balance in absorbers and strippers, Eq. (12-58)
ΔT_{lm}	log mean temperature difference for heat exchangers, Eq. (11-14)
ΔV_{feed}	change in vapor flow rate at feed stage at minimum reflux, Eqs. (7-28) and (7-29)
ε	Greek lowercase letter epsilon, limit for convergence
ε	porosity or void fraction, Eq. (14-4a)
ε_E	limit for energy convergence, Eq. (2-52)
ε_T	limit for temperature convergence, Eq. (6-19)
$\varepsilon_A, \varepsilon_B, \varepsilon_{AB}$	Lennard-Jones interaction energies, Table 15-2 and Eq. (15-21d)
ζ	Greek lowercase letter zeta, dimensionless distance, Eq. (15-14a)
η	Greek lowercase letter eta, fraction column cross-sectional area available for vapor flow, dimensionless
η	parameter in series solution for liquid falling-film mass transfer, Eq. (15-55b)
θ	Greek lowercase letter theta, angle of downcomer, Figure 10-19B
ι	Greek lowercase letter iota
κ	Greek lowercase letter kappa
λ	Greek lowercase letter lambda, latent heat of vaporization, kcal/kg, Btu/lb, Btu/lbmol, etc.
λ	$= \frac{m}{L/V}$ parameter for packed bed estimation of HETP, Eqs. (10-37) and (16-36)
μ	Greek lowercase letter mu, viscosity, cp or Pa · s = kg/(m·s)
μ_C, μ_D	viscosity of continuous and dispersed phases, Pa · s
μ_H, μ_L	viscosity of heavy and light phases, Pa · s
μ_L	liquid viscosity, cp, Eqs. (10-5) and (10-7)
μ_M, μ_{mix}	mixture viscosity, Eqs. (10-8a) and (13-48), cp or Pa · s
μ_w	viscosity of water, cp
ν	Greek lowercase letter nu, kinematic viscosity = μ/ρ , m ² /s, cm ² /s, ft ² /s
ξ	Greek lowercase letter xi, warped time, Eqs. (9-28a)
ρ	Greek lowercase letter rho, density, g/cm ³ , lb/ft ³ , or kg/m ³
ρ_C, ρ_D	densities of continuous and dispersed phases, kg/m ³
ρ_L	Greek lowercase letter rho, liquid density, g/cm ³ , lb/ft ³ , or kg/m ³
ρ_{Mix}	mixture density, Eq. (13-47), kg/m ³
ρ_{mol}	molar density, kmol/m ³
$\rho_{\text{S,mol}}$	solid molar density, kmol/m ³
ρ_V	vapor density, g/cm ³ , lb/ft ³ , or kg/m ³
$\Delta \rho$	$ \rho_C - \rho_D $

σ	Greek lowercase letter sigma, surface tension, dynes/cm or interfacial tension
Φ_c, Φ_d	Greek lowercase letter phi, volumetric fraction of continuous and dispersed phases
$\Phi_{d,feed}$ ϕ or φ	volumetric fraction of dispersed phase in feed, Eq. (13-44) Greek phi symbol, Greek letter phi, liquid-phase packing parameter, Eq. (16-38), Figure 16-6
$\phi = \frac{L_{min}}{V_{min} K_{ref}}$ and $\bar{\phi} = \frac{\bar{L}_{min}}{\bar{V}_{min} \bar{K}_{ref}}$	Underwood's variables enriching and stripping sections, dimensionless, Eq. (7-24)
ϕ_B ϕ_{dc}	solvent interaction parameter, Eq. (15-22b) relative froth density in downcomer, Eq. (10-29), average value = 0.50
χ	Greek lowercase letter chi, term defined in Eq. (13-45)
ψ	Greek lowercase letter psi, ρ_{water}/ρ_L (Chapter 10)
Ψ	$e/(e + L)$, fractional entrainment, Figure 10-16
Ψ	packing parameter for gas phase, Eq. (16-37) and Figure 16-5
Ω_D	Greek capital letter omega, collision integral, Eq. (15-21e) and Table 15-2
ω	Greek lowercase letter omega, revolutions per second or revolutions per minute

Chapters 17 and 18

a, b	empirical constants in Eq. (17-49)
A	kg/h anhydrous crystals
A	cumulative area/volume of crystals, m^2/m^3
A_c	cross sectional area of ingot in zone melting, m^2
A_C	crystal surface area for heat and mass transfer, m^2
A, B	constants in Margules Eq. (18-4) for binary activity coefficients
$A_{heat\ exchanger}$	area of heat exchanger, m^2
A_T	total area/volume of crystals, m^2/m^3
$B(L, t)$	birth function, Eq. (17-16)
B_o	rate of formation of nuclei, $\#/(s \cdot m^3)$, Eqs. (17-18a) and (17-18b)
c	concentration, kg/m^3 , g/L
c^*	equilibrium concentration of solute, kg/m^3 , g/L
c_I	interfacial concentration of solute, kg/m^3 , g/L
C	concentration, kg anhydrous crystals/kg water (Chapter 17)
C^*	equilibrium concentration, kg anhydrous crystals/kg water
C	flow rate crystals, kg/h, or in static systems amount of crystals, kg
C	deposition rate crystals in falling film, kg/s
\hat{C}	flow rate crystals, kmol/h or in static system amount of crystals, kmol
$C_{impurity,Cry, avg}$	concentration of impurity in crystals after sweating
$C_{impurity,melt}$	concentration in the melt before sweating
Const1	$= 1/h_{HTF} + 1/(k_{metal}/t_{wall}) + 1/h_{NC}$, Eq. (18-33a)
Const2	$= \left\{ \frac{(T_C - T_{Melt})}{\rho_{Cry} [\lambda_{freeze} + C_{P,Melt} (T_{Cry} - T_{Melt})]} \right\}$, Eq. (18-33d)
$C_{P,C}$	heat capacity at constant pressure for crystal, kJ/(kg K)
$C_{P,L}$	heat capacity at constant pressure for liquid, kJ/(kg K)
$C_{P,V}$	heat capacity at constant pressure for vapor, kJ/(kg K)

C_S	solid (ingot) concentration in zone melting, g impurity/g mixture, Figure 18-11
$C_3 = 1 + \frac{\rho_L}{\rho_C}$	grouping of terms in Eqs. (18-23)
d_h	hydraulic diameter, m, Eq. (18-32a)
d_i	impeller diameter, m
$D(L, t)$	death function, Eq. (17-16)
D	diffusivity, m^2/s
Dia	diameter of particle, m
e	entrainment, Eq. (18-5a), kg melt/kg crystals
e_v	volumetric entrainment, m^3 entrained melt/ $(m^3$ total crystals plus melt), Eq. (18-6a)
E	flow rate of entrained melt, kg/h
\hat{E}	flow rate of entrained melt, kmol/h, Eq. (18-6e)
E_{Sw}	proposed sweating efficiency, $E_{Sw} = \Delta k_{eff,Sw} / k_{eff,before}$
f	fraction crystallized in progressive freezing and zone melting
F	feed rate, kg/h, kg/s
\hat{F}	feed rate, kmol/h or in static system amount of feed, kmol
F_{HTF}	feed rate heat transfer fluid, kg/s
F_W	liquid water flow rate, kg/h
g	$9.81 m/s^2$, the acceleration due to gravity
G	linear growth rate, m/s
Gr	Grashoff number, Eq. (18-29a, b), dimensionless
h	liquid enthalpy, kJ/kg
h_C	heat transfer coefficient at crystal surface, $kW/(m^2 s)$
h_{cool}	heat transfer coefficient for HTF in static system, $kW/(m^2 s)$
h_{Cry}	crystal enthalpy, kJ/kg, Eq. (18-41e)
h_{ff}	heat transfer coefficient for falling film, $kW/(m^2 s)$
$h_{ff,HTF,avg}$	average falling film heat transfer coefficient for HTF, $kW/(m^2 s)$
$h_{ff,proc}$	falling film heat transfer coefficient on process side, $kW/(m^2 s)$
h_{HTF}	enthalpy heat transfer fluid, kJ/kg
h_{magma}	enthalpy of magma, kJ/kg
h_M	enthalpy of mixing point, kJ/kg
h_{NC}	natural convection heat transfer coefficient, $kW/(m^2 s)$
h_{sphere}	heat transfer coefficient for sphere, $kW/(m^2 s)$
H_V	vapor enthalpy, kJ/kg
HTF	heat transfer fluid
i	order of nucleation
k	empirical exponent, Eq. (17-18b)
k_A	area shape factor
k_{cond}	thermal conductivity, $W/(m K)$
k_C	thermal conductivity of crystal, $W/(m K)$
$k_{eff,imp}$	approximate inclusion of mass transfer in equilibrium analysis of zone refining, Eq. (18-58)
$k_{eff,Cry}$	effective impurity distribution coefficient, Eqs. (18-59a, b)
k_{imp}	linear equilibrium constant of impurity,
$k_{imp}(T)$	$= \frac{\text{Weight fraction of impurity } i \text{ in solid}}{\text{Weight fraction of impurity } i \text{ in liquid}} = \frac{x_{i,S}}{x_{i,L}}, \text{ Eq. (18-5b)}$
k_{metal}	thermal conductivity of metal wall of heat exchanger, $W/(m K)$
k_M	film mass transfer coefficient, m/s
k_N	empirical rate constant for nucleation, Eq. (17-18b)
k_r	rate constant

k_v	volume shape factor
K_{salt}	empirical constant, Eq. (17-52)
K_G	overall mass transfer coefficient, m/s
K_{imp}	$= x_{\text{interface,imp}}/x_{\text{C,imp}}$, impurity distribution coefficient
K_S	solubility product, Eq. (17-50a)
l	length of solid melted in zone melting, m
L, L_{flow}	liquid flow rate, kg/h, kg/s
\hat{L}	liquid flow rate, kmol/h or in static system amount liquid, kmol
L	characteristic dimension of crystal, m
L	length of plate, m
\bar{L}	average crystal length, m
L	cumulative length per volume, m/m^3
L_o	initial weight of liquid in dilute progressive freezing, kg
L	weight of liquid in dilute progressive freezing, kg
L_p	size product crystals, m,
L_s	size seed crystals, m
L_T	total length of particles/volume, m/m^3
m	mass of ingot in zone melting, kg
m_{Cry}	mass of crystal, kg
m	valence of cation, Eq. (17-50a)
M	mixed stream amount or flow rate, kg or kg/h
M	cumulative mass of crystals/volume, kg/m^3
M_o	mass of molten zone in zone melting, kg, Eq. (18-52b)
M_T	magma density, weight of crystals in solution per volume, kg/m^3
MW	molecular weight
n	population density, $\#/(m \cdot m^3)$, Eqs. (17-12a) and (17-12b)
n	empirical exponent, Eq. (17-18b)
n	overall growth rate order
n	valence of anion, Eq. (17-50a)
n	moles water per mole hydrate
n	number of zone melting passes, Table 18-9.
$n_{\text{anhyd} \rightarrow \text{hydrate}}$	moles anhydrous compound per mole hydrate = 1.0
n_o	population density of nuclei, $\#/(m \cdot m^3)$
n_r	order for growth
N	cumulative number crystals/volume
$N_{A,\text{mol}}, N_{B,\text{mol}}$	molar fluxes, $\text{kmol}/(\text{h m}^2)$
NC	natural convection
N_{Po}	power number. $P/(\rho_M \omega^3 d_i^5)$, dimensionless
N_{Re}	Reynolds number for particle suspension, Eq. (18-19b), based on Stokes velocity, dimensionless
N_s	stirrer speed, rpm
N_T	total number of particles/volume, $\#/\text{m}^3$
N_w	number of gram-formula weights of solute per 1000g of H_2O , Eqs. (18-2a) and (18-2b)
Nu	Nusselt number for heat transfer, Eqs. (18-30a), (18-32c), and (18-46c), dimensionless
P	power, kW
P	product flow rate, kg/s
Pr	Prandtl number, $Pr = (\mu C_p)/k_{\text{cond}}$, Eq. (18-29e), (18-32d), and (18-46c), dimensionless
Q	volumetric flow rate, m^3/s
Q_{cool}	cooling load, kJ/s

Q_{energy}	heat or cooling load, kJ/s
Q_{growth}	cooling load required for crystal growth, kJ/s, Eq. (18-27)
Q_{sensible}	cooling load required to cool wall, crystal and melt when T_{HTF} is reduced
R	ideal gas constant (see Appendix C for values)
Ra	Raleigh number = $Gr Pr$, Eq. (18-29c, d), dimensionless
Re	Reynolds number $(Dia)\nu\rho/\mu$, Eq. (18-18a), dimensionless
Re_{ff}	falling film Reynolds number = $4 \Gamma/\mu$, Eq. (18-45b)
Residue	flow rate of residue in suspension melt crystallization, kg/h
Sc	kg/h crystals (stable hydrate form), Eqs. (17-4e) and (17-4f)
Sc	Schmidt number, $Sc = \mu/(\rho_r D_{AB})$, Eqs. (18-18a) (18-31c), dimensionless
Sh	Sherwood number, $(k_M Dia/D_{AB})$, Eqs. (18-18a), (18-31a), dimensionless
t	time
t_C	thickness of crystal layer, m
t_P	thickness of heat transfer plate, m
T	temperature, °C or K
T^*	temperature at equilibrium, °C or K
T_C, T_{cold}	coolant temperature, °C or K
T_{Cry}, T_I	temperature of crystal surface, °C or K
T_E	eutectic temperature, °C or K
T_{Eq}	equilibrium temperature at x_1 , °C or K
$T_{\text{Eq,melt}}$	equilibrium temperature of melt before sweating, °C or K
T_M	magma temperature, °C or K
T_{melt}	melting point temperature, °C or K
T_{melt}	temperature of melt, °C or K
T_{P1}, T_{P2}	temperatures on two sides of heat transfer plate, side 1 is on HTF side and side 2 has crystal growth, °C or K
T_{SW}	sweating temperature, °C or K
U	overall heat transfer coefficient, Eq. (18-26c), kW/(m ² s)
v_T	particle Stokes' law terminal velocity, Eq. (18-19c), m/s
V	volume magma, m ³
V	vapor evaporation rate, kg/h
V	volume of crystal, m ³
V_{sample}	volume of sample of magma sieved, m ³
Vol	volume of crystals/volume, m ³ /m ³ (Chapter 17)
Vol_T	total volume of crystals /volume, m ³ /m ³
w	width of heat transfer plate, m
W_i	weight of crystals in sieve, m ³
\dot{W}_p	flow rate of product crystals, kg/h
\dot{W}_p	batch crystallization, kg product/kg initial water
\dot{W}_s	flow rate of seed crystals, kg/h
\dot{W}_s	batch crystallization, kg seed crystals/kg initial water
x	mass fraction solute in solid or in solution
x^*	mole fraction solute in solution at saturation, Eq. (17-2)
$x_{A,\text{surf,mol}}$	mole fraction desired product A in fluid at crystal surface
x_{bulk}	mass or mole fraction impurity in bulk fluid
x_C	mass or mole fraction impurity in crystal
$x_{C,\text{imp}}$	mass fraction impurity in crystal phase in zone melting and progressive freezing

$x_{C,imp,o}$	uniform initial mass fraction impurity in crystal phase in zone melting
$x_{C,salt}$	mass fraction salt in crystals (hydrates also contain water)
x_E	eutectic mass or mole fraction of impurity
x_I	mass or mole fraction impurity at crystal fluid interface.
$x_{L,imp}$	mass fraction impurity in liquid phase in zone melting and progressive freezing
$x_{L,imp,o}$	uniform initial mass fraction impurity in liquid phase in progressive freezing
$x_{L,salt}$	mass fraction salt in liquid
$x_{salt,mix}$	mass fraction salt at mixing point M
$x_{M,A}$ and $x_{A,mol}$	mole fraction acetic acid, Example 18-2 and Tables 18-3a and 3b
$x_{M,C}$	mole fraction solute (impurity) in crystals
$x_{M,F}$ and $x_{F,mol}$	mole fraction solute (impurity) in feed
$x_{M,L}$ and $x_{L,mol}$	mole fraction solute (impurity) in liquid
$x_{M,W}$ and $x_{W,mol}$	mole fraction water, Example 18-2 and Tables 18-3a, and 3b
y_{salt}	mass fraction salt in vapor, typically 0
z	distance along ingot in zone melting, m, Figure 18-10
z_{cut}	distance from starting end of ingot to where ingot is cut to harvest purified portion in zone melting, m, Figure 18-10d
Greek	
β	Greek lowercase letter beta, empirical constant, Eq. (17-52)
β	coefficient of volumetric expansion, Eq. (18-29e, f)
γ	Greek lowercase letter gamma, activity coefficient
Γ	Greek capital letter gamma, falling film vertical (L direction) mass rate of flow per second per meter in the w direction, kg/(s m)
$\Gamma_{crystal}$	reduction in Γ from crystal deposition on wall, kg/(s m), Eqs. (18-47b, c)
Γ_{feed}	feed rate to plate expressed as Γ , kg/(s m)
δ	Greek lowercase letter delta, falling film thickness, m, Eqs. (18-46d, e)
$\delta_{mass} = D/k_M$	mass transfer boundary layer or film thickness, m
$\delta_{thermal} = k_{cond}/h$	thermal boundary layer or film thickness, m
Δc	Greek capital letter delta, concentration supersaturation, $c - c^*$, kg/m ³
Δk_{eff}	$= k_{eff,before} - k_{eff,after}$ measure of change after sweating, Eq. (18-60a)
ΔL_{growth}	$L_p - L_s$, m
ΔL_i	size range sieve fraction i, m
Δp_i	pervaporation driving force, partial pressure difference, Eq. (19-49c)
Δt	freezing-point lowering in °C, Eq. (18-2b)
ΔW_i	weight of crystals collected on screen i, kg
ϵ	Greek lowercase letter epsilon, void fraction of crystal layer
η	Greek lowercase letter eta, empirical exponent, Eq. (17-18a)
λ_{freeze}	Greek lowercase letter lambda, molar heat of fusion, kJ/kmol
μ	Greek lowercase letter mu, viscosity, cp or (Pa s) = kg/(m s)
μm	micron, 10 ⁻⁶ m
Π_C	Greek capital pi, crystallization process dimensionless number = $G/k_{M,imp}$
Π_{Sw}	dimensionless sweating number, $\Pi_{Sw} = k_{eff,before} (T_{Sw} / T_{Eq,melt})$
ρ_C	Greek lowercase letter rho, crystal solid density, kg/m ³
$\rho_{C,mol}$	crystal solid density, kmol/m ³

$\rho, \rho_{\text{fluid}}$	fluid density, kg/m ³
$\hat{\rho}_{f,\text{entrained}}$	density of entrained fluid, kmol/m ³
ρ_F	feed density, kg/m ³
$\rho_{L,\text{mol}}$	liquid molar density, kmol/m ³
ρ_M	magma density, kg/m ³
ρ_{out}	suspension density out, kg/m ³
τ	Greek lowercase letter tau, $\tau = V/Q_{\text{out}}$, drawdown or retention time, s or h
ω	Greek lowercase letter omega, revolutions/s

Symbols

[...]	concentration in moles/L
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Chapter 19

a	term in quadratic equation for well-mixed membrane system, Eq. (19-10b)
a_m	constant in expression to calculate osmotic pressure, kPa/mole fraction, Eq. (19-14c), if van't Hoff equation is valid, $a_m = RT$, Eq. (19-14e)
a'	constant in expression to calculate osmotic pressure, kPa/weight fraction, Eq. (19-14d, h, i)
A, A_{total}	membrane area available for mass transfer, cm ² or m ²
A_i	membrane area of each stage in staged model, cm ² or m ²
b	term in quadratic equations for well-mixed membrane systems, Eq. (19-10c)
Barrer	$=1 \times 10^{-10} \text{ cm}^3(\text{STP})\text{cm}/[\text{cm}^2 \text{ s cm Hg}]$
c	term in quadratic equations for well-mixed membrane systems, Eq. (19-10d)
c, C	concentration, g solute/cm ³ solution or g solute/L solution or similar units
c_{out}	outlet concentration of solute, g/L
c_p	permeate concentration of solute, g/L
c_w	concentration of solute at wall, g/L
constant	constant in mass transfer correlation, dimensionless, Eqs. (19-37a, b)
$C_{i,\text{final}}$	final concentration in diafiltration, g/L
$C_{i,o}$	initial concentration in diafiltration, g/L
$C_{PL,p}$	liquid heat capacity of permeate, kJ/(kg·°C)
$C_{PV,p}$	vapor heat capacity of permeate, kJ/(kg·°C)
C_s	salt concentration, kg/m ³
d_t	diameter of tube, cm
d_{tank}	tank diameter, cm
D	diffusivity in solution, cm ² /s, m ² /s
$D_{A,M}$	diffusivity of solute A in the membrane, cm ² /s, m ² /s
D_{Brownian}	diffusivity due to Brownian motion, cm ² /s, m ² /s, Eq. (19-45a)
D_{Shear}	shear induced hydrodynamic diffusivity in MF, cm ² /s, m ² /s, Eq. (19-45b)
F_p	volumetric flow rate of permeate, cm ³ /s
F_{out}	volumetric flow rate of exiting retentate, cm ³ /s

F_{solv}	volumetric flow rate of solvent in RO, cm^3/s
F_{m}	molar flow rate, mol/s, mol/min, etc.
$F_{\text{m,in}}, F_{\text{m,p}}, F_{\text{m,r,out}}$	molar flow rates for feed, permeate, and retentate, mol/s, kmol/h
$F'_{\text{m}}, F'_{\text{m,in}}, F'_{\text{m,p}}, F'_{\text{m,r,out}}$	mass flow rates, g/s, g/min, kg/min, etc.
h	1/2 distance between parallel plates, cm
h_{in}	enthalpy of inlet liquid stream in pervaporation, kJ/kg
h_{out}	enthalpy of outlet liquid retentate stream in pervaporation, kJ/kg
H_A	solubility parameter, $\text{cc}(\text{STP})/[\text{cm}^3 \cdot (\text{cm Hg})]$
H_p	enthalpy of vapor permeate stream in pervaporation, kJ/kg
j	counter for stage location in staged models in Figure 19-20
J	volumetric flux, $\text{cm}^3/(\text{s} \cdot \text{cm}^2)$ or $\text{m}^3/(\text{m}^2 \cdot \text{day})$, Eq. (19-1b)
J'	mass flux, $\text{g}/(\text{s} \cdot \text{cm}^2)$ or $\text{g}/(\text{m}^2 \cdot \text{day})$, Eq. (19-1c)
$J'_{\text{pure water}}$	pure water mass flux, $\text{g}/(\text{s} \cdot \text{cm}^2)$ or $\text{g}/(\text{m}^2 \cdot \text{day})$, Eq. (19-28)
J'_{solv}	mass flux solvent in RO, $\text{g}/(\text{s} \cdot \text{cm}^2)$ or $\text{g}/(\text{m}^2 \cdot \text{day})$, Eq. (19-13a, b)
J'_{solv}	mass flux solvent in UF, $\text{g}/(\text{s} \cdot \text{cm}^2)$ or $\text{g}/(\text{m}^2 \cdot \text{day})$, Eq. (19-39)
$J'_{\text{solv,osmotic}}$	mass flux osmotic flow, $\Delta p = 0$, $\text{g}/(\text{m}^2 \cdot \text{s})$ Eq. (19-14a)
J_{m}	mole flux, $\text{mol}/(\text{s} \cdot \text{cm}^2)$ or $\text{kmol}/(\text{day} \cdot \text{m}^2)$, Eq. (19-1d)
k	mass transfer coefficient, typically cm/s, Eqs. (19-33) and (19-36b)
k	mass transfer coefficient due to shear, cm/s, Eq. (19-45c)
k_B	Boltzmann constant, see Appendix C
K'_A	RO solute permeability, $\text{g}/(\text{m} \cdot \text{s} \cdot \text{weight fraction})$, Eq. (19-15b)
$K_{\text{M},i}$	multicomponent gas permeation rate transfer term, dimensionless, Eq. (19-11c)
K'_{solv}	GP permeability of solvent through membrane, $\text{L}/(\text{atm} \cdot \text{m}^2 \cdot \text{day})$ or similar units
K'_{solv}	RO or UF solvent permeability, $\text{g}/(\text{m} \cdot \text{s} \cdot \text{atm})$, Eq. (19-13b) or Eq. (19-39)
K'_{water}	RO water permeability, $\text{g}/(\text{m} \cdot \text{s} \cdot \text{atm})$, Eq. (19-28)
L	tube length, cm
M	concentration polarization modulus, weight fraction units, dimensionless, Eqs. (19-16), (19-48c)
M_c	concentration polarization modulus, concentration units, dimensionless, Eq. (19-48c)
M_{inner}	internal concentration modulus in FO, Eq. (19-38b)
MW	molecular weight, g/mol or kg/kmol
N	number of well-mixed stages in models in Figure 19-20
p	pressure, Pa, kPa, atm, mm Hg, etc.
p_A	partial pressure of species A, Pa, atm, mm Hg, etc.
$p_{i,1}, p_{i,2}$	partial pressure component i on upstream and downstream sides, respectively in pervaporation
p_p	total pressure on the permeate (low pressure) side, Pa, kPa, atm, mm Hg, etc.
p_r	total pressure on the retentate (high-pressure) side, Pa, kPa, atm, mm Hg, etc.
P_A	permeability of species A in GP membrane, $[\text{cc}(\text{STP}) \cdot \text{cm}]/[\text{cm}^2 \cdot \text{s} \cdot \text{cm Hg}]$
Q	volumetric flow rate, cm^3/s
R, R_i	rejection coefficient, weight fraction units, dimensionless, Eqs. (19-24a) and (19-48a)
R_c	rejection coefficient, concentration units, dimensionless, Eq. (19-48a)
R°	inherent rejection coefficient ($M = 1.0$), dimensionless, Eq. (19-48b)
R°	in UF and MF fraction of flux carried by pores that exclude solute

$R^{\circ}c$	inherent rejection coefficient ($M_c = 1.0$), concentration units, dimensionless, Eq. (19-48b)
$R_{\text{mid-size}}$	rejection coefficient of particles that are partially rejected in diafiltration
R	tube radius, cm, m
R	gas constant, Appendix C
Re	Reynolds number, dimensionless, Eq. (19-35b)
Sc	Schmidt number, dimensionless, Eq. (19-35d)
Sh	Sherwood number, dimensionless, Eqs. (19-35a) and (19-35d)
STP	standard temperature and pressure, standard conditions in this text = 0°C and 101.3 bar
t	time, s, min, h
t_{ms}	thickness of membrane skin doing separation, μm , mm, cm, or m
T	temperature, °C, K
T_{ref}	reference temperature, °C, K
u_b	bulk velocity in tube, cm/s, m/s, ft/s
V_o	constant volume of protein solution in diafiltration, cm^3 , L, m^3 , etc.
V_{solvent}	partial molar volume of the solvent, cm^3/mol , L/mol, m^3/kmol , etc.
V_w	volumetric water consumption in diafiltration, cm^3 , L, m^3 , etc.
$(VP)_{i,1}$	vapor pressure component i on upstream side pervaporation, kPa
x	weight fraction of retentate in pervaporation; in binary system refers to more permeable species
x_g	weight fraction at which solute gels in UF and MF
x_{in}	weight fraction of solute entering RO, UF, and MF systems
x_{out}	weight fraction of solute in retentate product in RO, UF, and MF
x_p	weight fraction of solute in liquid permeate in RO, UF, and MF
x_r	weight fraction of solute in retentate in RO, UF, and MF
x_w	weight fraction of solute in retentate at membrane wall in RO, UF, and MF
y	weight fraction of permeate in pervaporation; in binary system refers to more permeable species
y_p	mole fraction of solute in gas permeate for gas permeation
y_r	mole fraction of solute in gas retentate for gas permeation
$y_{r,\text{out}}$	mole fraction of solute in retentate product for gas permeation
$y_{r,w}$	mole fraction of solute in gas retentate at membrane wall
z	distance from wall, cm, m, Figures 19-10 and 19-14

Greek letters

α	Greek lowercase letter alpha, selectivity, dimensionless; gas permeation: Eq. (19-4b), pervaporation: Eq. (19-60)
α'	selectivity in RO, Eq. (19-18)
$\beta_{\text{evaporation}}$	Greek lowercase letter beta, evaporation separation factor in pervaporation, dimensionless, Eq. (19-63a)
$\beta_{\text{pervaporation,A,B}}$	pervaporation selectivity, Eq. (19-50a, b)
$\beta_{\text{permeation}}$	permeation separation factor in pervaporation, dimensionless, Eq. (19-63b)
γ	Greek lowercase letter gamma, activity coefficient
γ_w	shear stress at wall = $4 u_b / (\text{tube Radius})$, 1/s
δ	Greek lowercase letter delta, film thickness, m, Figure 19-10
Δx	difference in weight fraction of solute across the membrane

$\Delta\pi$	difference in the osmotic pressure across the membrane, Pa, atm, mm Hg, etc.
π	Greek lowercase letter pi, osmotic pressure, Pa, kPa, atm, mm Hg, etc.
π_D	osmotic pressure of draw solution in FO, Pa, kPa, etc.
π_F	osmotic pressure of feed solution in FO, Pa, kPa, etc.
π	ratio circumference to diameter of circle, = 3.14159
θ	Greek lowercase letter theta, cut = F_p/F_{in} with flows in molar units, dimensionless
θ_j	cut = $F_p/F_{r,j-1}$ for each stage in staged model of GP, dimensionless
θ'	cut = F'_p/F'_{in} in flows in mass units, dimensionless
μ	Greek lowercase letter mu, viscosity, centipoise or g/(cm s)
$\nu = \mu/\rho$	Greek lowercase letter nu, kinematic viscosity, cm ² /s
ρ	mass density gas in GP, kg/m ³
ρ_m	molar density gas in GP, kmol/m ³
ρ_{solution}	mass density of solution, kg/m ³
ρ_{solv}	mass solvent density, kg/m ³
$\rho_{M,\text{solv}}$	molar solvent density, kmol/m ³
λ_p	Greek lowercase letter lambda, mass latent heat of vaporization of the permeate in pervaporation determined at the reference temperature, kJ/kg
ω	Greek lowercase letter omega, stirrer speed in radians/s

Chapters 20 and 21

a	constant in Langmuir isotherm, same units as q/c, Eq. (20-6c)
a	argument for error function, dimensionless, Eq. (21-24), Table 21-2
a_p	surface area of the particles per volume, m ⁻¹
A_c	cross-sectional area of column, m ²
A_w	wall surface area per volume of column for heat transfer, m ⁻¹
b	constant in Langmuir isotherm, (concentration) ⁻¹ , Eq. (20-6c)
c_A	concentration of species A, kg/m ³ , kmol/m ³ , g/L, etc.
c_i	concentration of species i, kg/m ³ , kmol/m ³ , g/L, etc., or
c_i	concentration of ion i in solution, typically equivalents/m ³
c_i^*	concentration of species i that would be in equilibrium with \bar{q}_i , same units as c_i
$\bar{c}_{i,\text{pore}}$	average concentration of solute i in pore, same units as c_i
$c_{i,\text{surface}}$	fluid concentration at surface of particles, $\epsilon_p = 0$, same units as c_i
c_{pore}	fluid concentration at surface of adsorbent pores, same units as c_i
c_{Ri}	concentration of ion i on the resin, typically equivalents/m ³
c_{RT}	total concentration of ions on the resin based on total column volume, typically equivalents/m ³
c_T	total concentration of ions in solution, typically equivalents/m ³
Const_i	constant relating solute i velocity to interstitial velocity, dimensionless, Eq. (20-15e)
$C_{P,f}$	heat capacity of the fluid, cal/(g·°C), cal/(mol·°C), J/(g K), etc.
$C_{P,p}$	heat capacity of particle including pore fluid, same units $C_{P,f}$
$C_{P,s}$	heat capacity of the solid, same units as $C_{P,f}$
$C_{P,w}$	heat capacity of the wall, same units as $C_{P,f}$
d_p	particle diameter, cm or m
D	desorbent rate in SMB, same units as F
D/F	desorbent to feed ratio in SMB, dimensionless

D_{col}	column diameter, m or cm
D_i	diffusivity including molecular and Knudsen diffusivities, m ² /s or cm ² /s, Eq. (21-4b)
D_{eff}	effective diffusivity, m ² /s or cm ² /s, Eqs. (20-4) and (21-3b)
D_K	Knudsen diffusivity, m ² /s or cm ² /s, Eq. (21-4a)
D_L	alternate expression for axial dispersion coefficient, m ² /s or cm ² /s, Eq. (21-15b)
D_m	molecular diffusivity in free solution, m ² /s or cm ² /s
D_s	surface diffusivity, m ² /s or cm ² /s, Eq. (21-5)
erf	error function, Eq. (21-24) and Table 21-2
E_D	axial dispersion coefficient due to both eddy and molecular effects, m ² /s or cm ² /s, determined from Chung and Wen Eq. (21-15a) and Eq. (21-14b) for N_{Pe}
E_{DT}	thermal axial dispersion coefficient, m ² /s or cm ² /s
E_{eff}	effective axial dispersion coefficient, same units E_D , Eq. (21-25)
f_{Cry}	fraction of adsorbent volume that is zeolite crystal in zeolite adsorbent
F	volumetric feed rate (e.g., m ³ /h, cm ³ /min, liter/h)
h_p	particle heat transfer coefficient, J/(K s m ²) or similar units
h_w	wall heat transfer coefficient, J/(K s m ²) or similar units
HETP	height of equilibrium plate, cm/plate, Eq. (21-32c)
k_f	film mass transfer coefficient, m/s or cm/s
$k_{m,c}$	lumped-parameter mass transfer coefficient with concentration driving force, m/s or cm/s, Eqs. (21-10b) and (21-12b)
$k_{m,q}$	lumped-parameter mass transfer coefficient with amount adsorbed driving force, m/s or cm/s, Eqs. (21-10a) and (21-12a)
K_A	adsorption equilibrium constant, $K_{A,c}$, or $K_{A,p}$
K_{AB}	mass action equilibrium constant for monovalent-monovalent ion exchange, dimensionless, Eq. (20-40a)
$K_{A,c}$	adsorption equilibrium constant in terms of concentration, (concentration) ⁻¹
$K'_{i,c}$	linearized adsorption equilibrium constant, concentration units, same units as q/c, Eq. (20-6b)
$K_{A,o}$	pre-exponential factor in Arrhenius Eq. (20-7a), same units as K_A
$K_{A,p}$	adsorption equilibrium constant in terms of partial pressure, (pressure) ⁻¹
$K'_{A,p}$	linearized adsorption equilibrium constant in terms of partial pressure, Eq. (20-5b)
K_d	size exclusion parameter, dimensionless
K_{DB}	mass action equilibrium constant for divalent-monovalent ion exchange, Eq. (20-41)
K_{DE}	Donnan exclusion factor, dimensionless, Eq. (20-44)
L	length of packing in column, m or cm
L_{MTZ}	length of mass transfer zone, Figure 21-3, m or cm
M	molecular weight of solute, g/mol or kg/kmol
N_i	equivalent number of plates in chromatography for solute i, Eq. (21-32a)
N_{Pe}	Peclet number based on particle diameter, dimensionless, Eq. (21-14b)
p_A	partial pressure of species A, mm Hg, kPa, etc.
p_{after}, p_{before}	pressures in PSA after and before pressure change
p_h	high pressure, mm Hg, kPa, etc.

P_L	low pressure, mm Hg, kPa, etc.
$Pe_{L,i}$	Peclet number based on length for solute i, dimensionless, Eq. (21-32b)
q_A	amount of species A adsorbed, kg/kg adsorbent, mol/kg adsorbent, or kg/L
$q_{A,max}$	maximum amount of species A that can adsorb, kg/kg adsorbent, mol/kg adsorbent, or kg/L
q_F	amount adsorbed in equilibrium with feed concentration, same units as q_A
\bar{q}_i	average amount of species i adsorbed, kg/kg adsorbent, mol/kg adsorbent, or kg/L
q_i^*	amount adsorbed that would be in equilibrium with fluid of concentration c_i , same units as q_A
Q	volumetric flow rate, m^3/s , L/min, etc.
r_p	pore radius, m or cm
R	resolution, dimensionless, Eq. (21-36)
R	gas constant (e.g., $R = 8.314 \frac{m^3 Pa}{mol K}$), see Appendix C for additional values)
Re	Reynolds number, dimensionless, Eq. (21-14b)
Sc	Schmidt number, dimensionless, Eq. (21-14b)
Sh	Sherwood number, dimensionless, Eq. (21-14b)
t	time, s, min, or h
t_{br}	breakthrough time, s, min, or h
t_{center}	time center of pattern exits column, s, min, or h, Eq. (21-46b)
$t_{elution}$	elution time, s, min, or h
t_F, t_{feed}	feed time, s, min, or h
t_{MTZ}	time of mass transfer zone, Figure 21-3, s, min, or h
t_R	retention time, s, min, or h
t_{sw}	switching time in SMB, s, min, or h
T	temperature, °C or K
T_{amb}	ambient temperature, °C or K
T_S	solid temperature, °C or K
T_W	wall temperature, °C or K
$u_{ion,i}$	velocity of ion i, m/s or cm/s
u_S	average solute velocity, m/s or cm/s
\bar{u}_S	average of solute velocities for A and B, cm/s, Eq. (21-37)
$u_{S,ion,i}$	diffuse wave velocity of ion i, m/s or cm/s
u_{sh}	shock wave velocity, m/s or cm/s
$u_{sh,ion,i}$	shock wave velocity of ion i, m/s or cm/s
u_{th}	thermal wave velocity, m/s or cm/s
u_{total_ion}	velocity of total ion wave, m/s or cm/s
$v_{A,product}$	interstitial velocity of A product if it was in the column, $m/s = (m^3/s \text{ A product})/(\epsilon_e A_c)$
$v_{B,product}$	interstitial velocity of B product if it was in the column, $m/s = (m^3/s \text{ B product})/(\epsilon_e A_c)$
v_D	interstitial velocity of desorbent if it was in the column, $m/s = (m^3/s \text{ D product})/(\epsilon_e A_c)$
v_{Feed}	interstitial velocity of feed if it was in the column, $m/s = (m^3/s \text{ feed})/(\epsilon_e A_c)$
v_{inter}	interstitial velocity, m/s or cm/s, Eq. (20-2b)
v_{super}	superficial velocity, m/s or cm/s, Eq. (20-2a)
$V_{available}$	volume available to molecule, m^3 , Eq. (20-1c)

V_{col}	column volume, m^3
V_{feed}	volume feed gas, m^3
V_{fluid}	volume available to fluid, m^3 , Eq. (20-1a)
V_{purge}	volume purge gas, m^3
W_A, W_B	width of chromatographic peak, s, min, or h
W	weight of the column per length, kg/m
x	deviation from the location of the peak maximum, dimensionless, Eq. (21-33)
x_l	deviation from peak maximum in length units, Eq. (21-34b)
x_t	deviation from peak maximum in time units, Eq. (21-34a)
x	weight or mole fraction solute in liquid, $\text{kg solute/kg liquid}$ or $\text{kmol solute/kmol liquid}$, dimensionless
x_i	$= c_i/c_T$ equivalent fraction of ion in solution, dimensionless
$X_{\text{breakthrough}}(z,t)$	general solution for column breakthrough for linear isotherms, same units as c , Eq. (21-26)
y	weight or mole fraction solute in gas, kg solute/kg gas , or $\text{kmol solute/kmol gas}$, dimensionless
$Y_{A,\text{after}}, Y_{A,\text{before}}$	mole fraction solute A in gas after and before PSA pressure change, Eq. (20-28a)
y_i	$= c_{Ri}/c_{RT}$ equivalent fraction of ion on resin, dimensionless
z	axial distance in column, m or cm (Measured from closed end for PSA calculations)
$Z_{\text{after}}, Z_{\text{before}}$	location of solute waves after and before PSA pressure change, Eq. (20-28b)

Greek letters

β_{strong}	ratio velocities of strong and weak solutes, Eq. (20-27), dimensionless
Δc	change in solute concentration, same units as c
ΔH_{ads}	heat of adsorption, J/kg , cal/mol , etc.
Δp_A	change in partial pressure, kPa , atm , etc.
Δq	change in amount adsorbed, kmol/kg adsorbent , kg/kg adsorbent , kmol/m^3 , or kg/m^3
Δt	change in time, s, min, or h
ΔT_f	change in fluid temperature, $^{\circ}\text{C}$ or K
Δz	increment of column length, m
γ	volumetric purge to feed ratio in PSA, dimensionless, Eq. (20-26)
γ_1, γ_2	constants in Eq. (21-15b), $\gamma_1 \approx 0.7$, $\gamma_2 \approx 0.5$ for $d_p > 0.002$ mm
ϵ_e	external porosity, dimensionless
ϵ_p	internal or pore porosity, dimensionless
ϵ_T	total porosity, dimensionless, Eq. (20-1b)
ρ_b	bulk density of adsorbent, kg/m^3 , Eq. (20-3b)
ρ_f	fluid density, kg/m^3
$\rho_{m,f}$	molar density of fluid, kmol/m^3
ρ_p	particle density, kg/m^3 , Eq. (20-3a)
ρ_S	structural density of solid, kg/m^3
σ	standard deviation of Gaussian chromatographic peak, Eq. (21-33)
σ_l	standard deviation in length units, m or cm, Eq. (21-34b)
σ_t	standard deviation in time units, min or s, Eq. (21-34a)
τ	tortuosity, dimensionless, Eq. (20-4)
τ	$\tau = t - z/u_{sh}$ Convert PDE for constant pattern to ODE, Eq. (21-39)
ζ	Greek letter zeta used as dummy variable in Eq. (21-24)

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Approximate Shortcut Methods for Multicomponent Distillation

Chapters 5 and 6 introduced multicomponent distillation. Matrix methods are efficient, but they still require a fair amount of time even on a fast computer. In addition, they are simulation methods and require a known number of stages and a specified feed plate location. Fairly rapid approximate methods are required for preliminary economic estimates, for recycle calculations where the distillation is only a small portion of the entire system, for calculations for control systems, and as a first estimate for more detailed simulation calculations.

In this chapter, we first develop the Fenske equation, which allows calculation of multicomponent separation at total reflux. Then we switch to the Underwood equations, which allow us to calculate the minimum reflux ratio. To predict the approximate number of equilibrium stages, we then use the empirical Gilliland correlation that relates the actual number of stages to the number of stages at total reflux, the minimum reflux ratio, and the actual reflux ratio. The feed location can also be approximated from an empirical correlation.

7.0 SUMMARY—OBJECTIVES

In this chapter, we develop approximate shortcut methods for binary and multicomponent distillation. After completing this chapter, you should be able to satisfy the following objectives:

1. Derive the Fenske equation, and use it to determine the number of stages required at total reflux and the splits of non-key (NK) components
2. Use the Underwood equations to determine the minimum reflux ratio for multicomponent distillation
3. Use the Gilliland correlation to estimate the actual number of stages in a column and the optimum feed stage location

7.1 TOTAL REFLUX: FENSKE EQUATION

Fenske (1932) derived a rigorous solution for binary and multicomponent distillation at total reflux. The derivation assumes that the stages are equilibrium stages. A multicomponent distillation column with a total condenser and a partial reboiler operating at total reflux is shown in Figure 7-1. For any two components A and B, equilibrium in the partial reboiler R requires

$$(y_A/y_B)_R = \alpha_{AB,R} \cdot (x_A/x_B)_R \quad (7-1)$$

Equation (7-1) is the definition of the relative volatility, Eq. (2-6a), applied to the reboiler. Material balances for components A and B around the reboiler are

$$V_R y_{A,R} = L_N x_{A,N} - B x_{A,R} \quad (7-2)$$

and

$$V_R y_{B,R} = L_N x_{B,N} - B x_{B,R} \quad (7-3)$$

However, at total reflux, $B = 0$, and $L_N = V_R$. Thus, the mass balances become

$$y_{A,R} = x_{A,N}, \quad y_{B,R} = x_{B,N} \quad (\text{total reflux}) \quad (7-4a, b)$$

For a binary system, this naturally means that the operating line is the $y = x$ line. Combining Eqs. (7-1) and (7-4),

$$\left[\frac{x_A}{x_B} \right]_N = \alpha_{AB,R} \left[\frac{x_A}{x_B} \right]_R \quad (7-5)$$

If we now move up the column to stage N , the equilibrium equation is identical to Eq. (7-1), except it is written for stage N . The mass balances around stage N simplify to $y_{A,N} = x_{A,N-1}$ and $y_{B,N} = x_{B,N-1}$. Combining these equations, we have

$$\left[\frac{x_A}{x_B} \right]_{N-1} = \alpha_{AB,N} \left[\frac{x_A}{x_B} \right]_N \quad (7-6)$$

Equations (7-5) and (7-6) can be combined to give

$$\left[\frac{x_A}{x_B} \right]_{N-1} = \alpha_{AB,N} \alpha_{AB,R} \left[\frac{x_A}{x_B} \right]_R \quad (7-7)$$

which relates the ratio of liquid mole fractions leaving stage $N-1$ to the ratio in the reboiler.

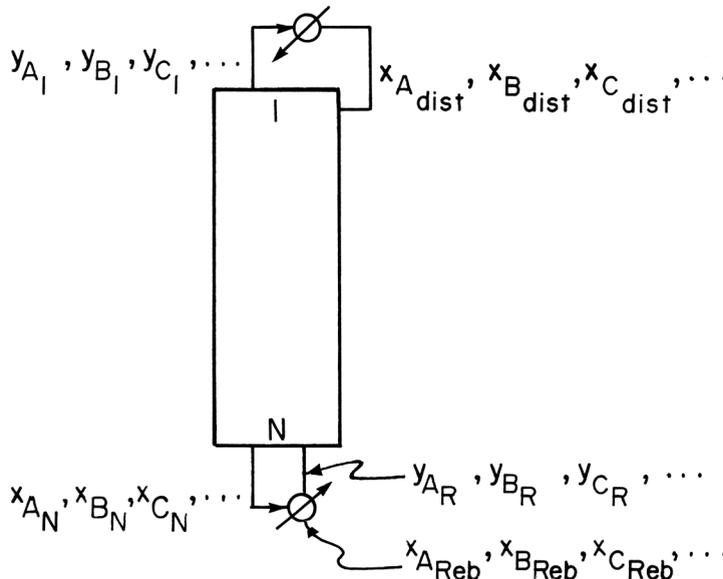


FIGURE 7-1. Total reflux column

Repeating this procedure of alternating between the operating and equilibrium equations, the result at the top stage is

$$\left[\frac{x_A}{x_B} \right]_{\text{dist}} = \alpha_{AB,1} \alpha_{AB,2} \alpha_{AB,3} \dots \alpha_{AB,N-1} \alpha_{AB,N} \alpha_{AB,R} \left[\frac{x_A}{x_B} \right]_{\text{R}} \quad (7-8a)$$

If we define α_{AB} as the geometric average relative volatility,

$$\alpha_{AB} = [\alpha_{AB,1} \alpha_{AB,2} \alpha_{AB,3} \dots \alpha_{AB,N-1} \alpha_{AB,N} \alpha_{AB,R}]^{1/N_{\min}} \quad (7-8b)$$

and note that reboiler and bottoms compositions are identical, Eq. (7-8a) becomes

$$\left[\frac{x_A}{x_B} \right]_{\text{dist}} = \alpha_{AB}^{N_{\min}} \left[\frac{x_A}{x_B} \right]_{\text{bot}} \quad (7-9)$$

Solving Eq. (7-9) for N_{\min} , we obtain

$$N_{\min} = \frac{\ln \left[\left(\frac{x_A}{x_B} \right)_{\text{dist}} / \left(\frac{x_A}{x_B} \right)_{\text{bot}} \right]}{\ln \alpha_{A-B}} \quad (7-10)$$

which is one form of the Fenske equation. N_{\min} is the number of equilibrium contacts including the partial reboiler required at total reflux. Although the derivation is for any two components, Eq. (7-10) is most accurate if it is written for the light key and the heavy key. Then we have

$$N_{\min} = \frac{\ln \left[\left(\frac{x_{LK}}{x_{HK}} \right)_{\text{dist}} / \left(\frac{x_{LK}}{x_{HK}} \right)_{\text{bot}} \right]}{\ln \alpha_{LK-HK}} \quad (7-11a)$$

If the relative volatility is constant, Eq. (7-10) is exact.

An alternative form of the Fenske equation that is very convenient for multicomponent calculations is easily derived. First, rewrite Eq. (7-11a) as

$$N_{\min} = \frac{\ln \left[\left(\frac{Dx_A}{Dx_B} \right)_{\text{dist}} / \left(\frac{Bx_A}{Bx_B} \right)_{\text{bot}} \right]}{\ln \alpha_{A-B}} \quad (7-11b)$$

$Dx_{A,\text{dist}}$ is equal to the fractional recovery of A in the distillate multiplied by the amount of A in the feed, and $Bx_{B,\text{bot}}$ is the fractional recovery of B in the bottoms multiplied by z_B ,

$$Dx_{A,\text{dist}} = (FR_{A,\text{dist}})(Fz_A) \quad \text{and} \quad Bx_{B,\text{bot}} = (FR_{B,\text{bot}})(Fz_B) \quad (7-11c, d)$$

These equations are the multicomponent equivalent of Eqs. (3-5a) and (3-5c). Substituting these equations and the equations for $Dx_{B,\text{dist}}$ and $Bx_{A,\text{bot}}$ into Eq. (7-11b), and identifying $A = LK$ and $B = HK$,

$$N_{\min} = \frac{\ln \left[\left(\frac{FR_{LK,\text{dist}}}{1 - FR_{LK,\text{dist}}} \right) / \left(\frac{1 - FR_{HK,\text{bot}}}{FR_{HK,\text{bot}}} \right) \right]}{\ln \alpha_{LK-HK}} \quad (7-12)$$

Equation (7-12) is in a convenient form for determining the number of stages for multicomponent systems.

For multicomponent systems, calculation with the Fenske equation is straightforward if fractional recoveries of the two keys, LK and HK, are specified. Equation (7-12) can be used directly to find N_{\min} . The relative volatility can be approximated by a geometric average. Once N_{\min} is known, the fractional recoveries of the NKs can be found by first writing Eq. (7-12) for an NK component and one of the key components. For example, if we replace LK with an NK component,

$$N_{\min} = \frac{\ln \left[\left(\frac{FR_{NK,dist}}{1 - FR_{NK,dist}} \right) / \left(\frac{1 - FR_{HK,bot}}{FR_{HK,bot}} \right) \right]}{\ln \alpha_{NK-HK}}$$

Solving for $(FR_{NK,dist})$ the result is,

$$FR_{NK,dist} = \frac{(\alpha_{NK-HK})^{N_{\min}}}{\frac{FR_{HK,bot}}{1 - FR_{HK,bot}} + (\alpha_{NK-HK})^{N_{\min}}} \quad (7-13)$$

Remember that the order of subscripts on α_{AB} is important.

If two mole fractions are specified, say $x_{LK,bot}$ and $x_{HK,dist}$, the multicomponent calculation is more difficult. We cannot use the Fenske equation directly, but several alternatives are possible. If we can assume that all NKs are nondistributing, we can use the strategy used in Chapter 5 to do mass balances. Assume the NKs follow Eqs. (5-6a) to (5-8) and then calculate D and B from the summation equations, Eqs. (3-6a) and (3-6b). Once all distillate and bottoms compositions or values for $Dx_{D,i}$ and $Bx_{B,i}$ have been found, Eq. (7-11a) or (7-11b) can be used to find N_{\min} . Use the key components for this calculation. The assumption of nondistribution of the NKs can be checked with Eq. (7-13). If the original assumption is invalid, the calculated value of N_{\min} obtained for key components can be used to calculate the light non-key (LNK) and heavy non-key (HNK) compositions in distillate and bottoms. Then Eq. (7-11a) or (7-11b) is used again to obtain a more accurate estimate of N_{\min} .

If NKs distribute, a reasonable first guess for the distribution is required. This guess can be obtained by assuming that the distribution of NKs is the same at total reflux as it is at minimum reflux. The distribution at minimum reflux can be obtained from the Underwood equation (Case C) and is covered later.

The derivation up to this point has been for any number of components. If we now restrict ourselves to a binary system where $x_b = 1 - x_A = 1 - x$, and Eq. (7-11a) becomes

$$N_{\min} = \frac{\ln \left[\left(\frac{x}{1-x} \right)_{dist} / \left(\frac{x}{1-x} \right)_{bot} \right]}{\ln \alpha_{A-B}} \quad (7-14)$$

where $x = x_A$ is the mole fraction of the more volatile component (MVC). The use of the Fenske equation for binary systems is quite straightforward. With distillate and bottoms mole fractions of the MVC specified, N_{\min} is easily calculated if α_{AB} is known. If the relative volatility is not constant, α_{AB} can be estimated from a geometric average as shown in Eq. (7-8b). This can be estimated for a first trial as

$$\alpha_{AB} = (\alpha_{AB,dist} \times \alpha_{AB,R})^{1/2} \quad (7-15)$$

where $\alpha_{AB,R}$ is determined from the bottoms composition and $\alpha_{AB,dist}$ from the distillate composition.

Accurate use of the Fenske equation obviously requires an accurate value for the relative volatility. Smith (1963) covers in detail a method of calculating α by estimating temperatures and calculating the geometric average relative volatility. For approximate estimates this extra work is seldom necessary.

EXAMPLE 7-1. Fenske equation

A distillation column with a partial reboiler and a total condenser is separating a saturated vapor feed that is 40.0 mol% benzene (B), 30.0 mol% toluene (T), and 30.0 mol% cumene (C). Recovery of toluene in the distillate is 95%, and recovery of cumene in the bottoms is 98%. Reflux is a saturated liquid, and constant molal overflow (CMO) is valid. Pressure is at 1.0 atm. Relative volatilities are constant. Choosing toluene as the reference component, $\alpha_{B-T} = 2.25$ and $\alpha_{C-T} = 0.21$. Find the number of equilibrium stages required at total reflux, the recovery fraction of benzene in the distillate and in the bottoms, and the mole fractions in the distillate and bottoms.

Solution

- A. Define. A total reflux column was shown in Figure 7-1. For T = toluene (LK), C = cumene (HK), B = benzene (LNK), we have $\alpha_{BT} = 2.25$, $\alpha_{TT} = 1.0$, $\alpha_{CT} = 0.21$, which means $\alpha_{TC} = 1/0.21$. $z_T = 0.3$, $z_B = 0.4$, $z_C = 0.3$, $FR_{LK,dist} = FR_{T,dist} = 0.95$, and $FR_{HK,bot} = FR_{C,bot} = 0.98$.
- Find N at total reflux.
 - Find $FR_{B,dist}$ at total reflux.
 - Find mole fractions of distillate and bottoms at total reflux.
- B. Explore. Because operation is at total reflux and relative volatilities are constant, we can use the Fenske equation.
- C. Plan. Calculate N_{min} from Eq. (7-12) and then calculate $FR_{B,dist}$ from Eq. (7-13).
- D. Do it.
- Equation (7-12) gives

$$N_{min} = \frac{\ln \left[\left(\frac{FR_{LK,dist}}{1 - FR_{LK,dist}} \right) / \frac{1 - FR_{HK,bot}}{FR_{HK,bot}} \right]}{\ln \alpha_{LK-HK}} = \frac{\ln \left[\frac{0.95 / 0.05}{0.98 / 0.02} \right]}{\ln(1 / 0.21)} = 4.38$$

Note that $\alpha_{LK-HK} = \alpha_{tol-cumene} = 1/\alpha_{Cumene-tol} = 4.762$.

- Equation (7-13) gives

$$FR_{B,dist} = \frac{(\alpha_{B-HK})^{N_{min}}}{\frac{FR_{HK,bot}}{1 - FR_{HK,bot}} + (\alpha_{B-HK})^{N_{min}}} = \frac{(2.25 / 0.21)^{4.38}}{\frac{0.95}{0.05} + \left(\frac{2.25}{0.21} \right)^{4.38}} = 0.9985$$

Benzene recovery in bottoms = $1 - FR_{B,dist} = 0.0015$. Note that

$$\alpha_{B-HK} = \frac{K_B}{K_{HK}} = \frac{K_B / K_T}{K_{HK} / K_T} = \frac{\alpha_{BT}}{\alpha_{CT}} = \frac{\alpha_{benz-tol}}{\alpha_{cumene-tol}} = \frac{2.25}{0.21}$$

- $DX_{i,dist} = (FR_{i,dist})(Fz_i) = 0.99805(0.4F) = 0.3994F$.
 $D = \sum(DX_{i,dist}) = 0.9985(0.4F) + 0.95(0.3F) + (1 - 0.98)0.3F = 0.6904F$.
Then, $x_{Ben,dist} = 0.3994F/0.6904F = 0.5785$, $x_{Tol,dist} = 0.4128$, $x_{Cum,dist} = 0.0087$.
 $B = F - D = 1 - 0.6904F = 0.3096F$.
 $x_{Ben,bot} = (1 - 0.9985)(0.4F)/0.3096F = 0.001938$, $x_{Tol,bot} = 0.0485$, $x_{Cum,bot} = 0.9496$.
- E. Check. The results can be checked by calculating $FR_{C,dist}$ using component A instead of B. The same answer is obtained.
- F. Generalize. High recovery of a compound (e.g., the HK) in the bottoms means there will be very little of that compound in the distillate. Thus, the distillate is pure. To have high purity of the bottoms, we must have high recovery of the LK in the distillate.

7.2 MINIMUM REFLUX: UNDERWOOD EQUATIONS

For binary systems, the pinch point usually occurs at the feed plate. When this occurs, an analytical solution for the limiting flows can be derived (King, 1980) that is also valid for multicomponent systems as long as the pinch point occurs at the feed stage. However, multicomponent systems with nondistributing components will have separate pinch points in both the stripping and the enriching sections. If there are HNKs and/or LNKs, there will be nondistributing components unless the separation is sloppy, the NKs have volatilities that are very close to the keys, or a sandwich component is present. With nondistributing components, an analysis procedure developed by Underwood (1948) is used to find the minimum reflux ratio.

The development of the Underwood equations is quite complex and is presented in detail by Underwood (1948), Smith (1963), and King (1980). For most practicing engineers the details of the development are not as important as the use of the Underwood equations; we therefore follow the approximate derivation of Thompson (1981). Thus, we outline the important points but ignore the mathematical details of the derivation.

If there are nondistributing HNKs present, a pinch point of constant composition will occur at minimum reflux in the enriching section above where the HNKs are fractionated out. With nondistributing LNKs present, a pinch point will occur in the stripping section. For the enriching section in Figure 7-2, the mass balance for component i is

$$V_{\min} y_{i,j+1} = L_{\min} x_{i,j} + Dx_{i,\text{dist}} \quad (7-16)$$

At the pinch point, where compositions are constant,

$$x_{i,j-1} = x_{i,j} = x_{i,j+1}, \text{ and } y_{i,j-1} = y_{i,j} = y_{i,j+1} \quad (7-17)$$

The equilibrium expression can be written in terms of K values as

$$y_{i,j+1} = K_i x_{i,j+1} \quad (7-18)$$

Combining Eqs. (7-16) to (7-18) we obtain a simplified balance valid in the region of constant compositions.

$$V_{\min} y_{i,j+1} = L_{\min} y_{i,j+1} / K_i + Dx_{i,\text{dist}} \quad (7-19)$$

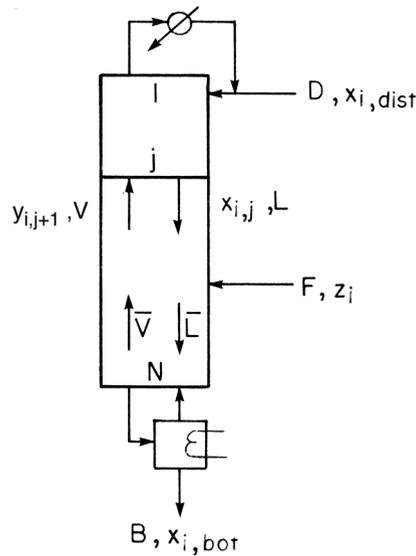


FIGURE 7-2. Distillation column

Defining the relative volatility $\alpha_{i\text{-ref}} = K_i/K_{\text{ref}}$ and combining terms in Eq. (7-19),

$$V_{\min} y_{i,j+1} [1 - L_{\min} / (\alpha_{i\text{-ref}} V_{\min} K_{\text{ref}})] = D x_{i,\text{dist}} \quad (7-20)$$

Solving for the component vapor flow rate, $V_{\min} y_{i,j+1}$, and rearranging

$$V_{\min} y_{i,j+1} = \frac{\alpha_{i\text{-ref}} D x_{i,\text{dist}}}{\alpha_{i\text{-ref}} - L_{\min} / (V_{\min} K_{\text{ref}})} \quad (7-21)$$

Equation (7-21) can be summed over all components to give the total vapor flow rate in the enriching section at minimum reflux:

$$V_{\min} = \sum_{i=1}^C (V_{\min} y_{i,j+1}) = \sum_{i=1}^C \left(\frac{\alpha_{i\text{-ref}} D x_{i,\text{dist}}}{\alpha_{i\text{-ref}} - L_{\min} / (V_{\min} K_{\text{ref}})} \right) \quad (7-22)$$

In the stripping section, a similar analysis can be used to derive

$$-\bar{V}_{\min} = \sum_{i=1}^C \left(\frac{\bar{\alpha}_{i\text{-ref}} B x_{i,\text{bot}}}{\bar{\alpha}_{i\text{-ref}} - \bar{L}_{\min} / (\bar{V}_{\min} \bar{K}_{\text{ref}})} \right) \quad (7-23)$$

Because the conditions in the stripping section are different than in the rectifying section, in general $\alpha_{i\text{-ref}} \neq \bar{\alpha}_{i\text{-ref}}$ and $K_{\text{ref}} \neq \bar{K}_{\text{ref}}$.

Underwood (1948) describes generalized forms of Eqs. (7-22) and (7-23) that are equivalent to defining

$$\phi = \frac{L_{\min}}{V_{\min} K_{\text{ref}}} \quad \text{and} \quad \bar{\phi} = \frac{\bar{L}_{\min}}{\bar{V}_{\min} \bar{K}_{\text{ref}}} \quad (7-24)$$

Equations (7-22) and (7-23) then become polynomials in ϕ and $\bar{\phi}$ with C roots. The equations are now

$$V_{\min} = \sum_{i=1}^c \frac{\alpha_{i\text{-ref}} (D x_{D,i})}{\alpha_{i\text{-ref}} - \phi} \quad \text{and} \quad -\bar{V}_{\min} = \sum_{i=1}^c \frac{\bar{\alpha}_{i\text{-ref}} (B x_{B,i})}{\bar{\alpha}_{i\text{-ref}} - \bar{\phi}} \quad (7-25a, b)$$

If we assume CMO and constant relative volatilities $\alpha_{i\text{-ref}} = \bar{\alpha}_{i\text{-ref}}$, Underwood showed there are common values of ϕ and $\bar{\phi}$ that satisfy both equations. Equations (7-25a) and (7-25b) can now be added. Thus, at minimum reflux,

$$V_{\min} - \bar{V}_{\min} = \sum_{i=1}^c \left[\frac{\alpha_{i\text{-ref}} (D x_{D,i})}{\alpha_{i\text{-ref}} - \phi} + \frac{\alpha_{i\text{-ref}} (B x_{B,i})}{\alpha_{i\text{-ref}} - \phi} \right] \quad (7-26)$$

where α is now an average volatility.

Eq. (7-26) is simplified with the external column mass balance

$$F z_i = D x_{D,i} + B x_{B,i} \quad (7-27)$$

to

$$\Delta V_{\text{feed}} = V_{\min} - \bar{V}_{\min} = \sum_{i=1}^c \frac{\alpha_{i\text{-ref}} (F z_i)}{\alpha_{i\text{-ref}} - \phi} \quad (7-28)$$

ΔV_{feed} is the change in vapor flow rate at the feed stage. If q is known,

$$\Delta V_{\text{feed}} = F(1 - q) \quad (7-29)$$

If the feed temperature is specified, a flash calculation on the feed can be used to determine ΔV_{feed} .

Equation (7-28) is known as the first Underwood equation. It can be used to calculate appropriate values of ϕ . Equation (7-25a) is known as the second Underwood equation and is used to calculate V_{min} . The exact method for using the Underwood equations depends on what can be assumed. We consider three cases.

Case A. Assume that none of the NKs distribute. In this case, the amounts of NKs in the distillate are

$$Dx_{D,\text{Hnk}} = 0 \text{ and } Dx_{D,\text{LNK}} = Fz_{\text{LNK}} \quad (7-30a, b)$$

while the amounts of the keys are

$$Dx_{D,\text{LK}} = (FR_{D,\text{LK}})Fz_{\text{LK}}, \text{ and } Dx_{D,\text{HK}} = [1 - (FR_{B,\text{HK}})]Fz_{\text{HK}} \quad (7-30c, d)$$

Solve Eq. (7-28) for the one value of ϕ between the relative volatilities of the two keys, $\alpha_{\text{HK-ref}} < \phi < \alpha_{\text{LK-ref}}$. This value of ϕ can be substituted into Eq. (7-25a) to calculate V_{min} . Then

$$D = \sum_{i=1}^c (Dx_{D,i}) \quad (7-31)$$

and L_{min} is found from mass balance

$$L_{\text{min}} = V_{\text{min}} - D \quad (7-32)$$

Case B. Assume that the distributions of NKs determined from the Fenske equation at total reflux are also valid at minimum reflux. In this case, the $Dx_{D,\text{LNK}}$ values are obtained from the Fenske equation as described earlier. Again solve Eq. (7-28) for the ϕ value between the relative volatilities of the two keys. This ϕ , the Fenske values of $Dx_{\text{NK,dist}}$, and the $Dx_{D,\text{LK}}$ and $Dx_{D,\text{HK}}$ values from Eqs. (7-30c) and (7-30d) are used in Eq. (7-25a) to find V_{min} . Then Eqs. (7-31) and (7-32) are used to calculate D and L_{min} . This procedure is illustrated in Example 7-2. Case C results are probably more accurate.

Case C. This case presents the exact solution without further assumptions. Equation (7-29) is a polynomial with C roots. Solve this equation for all values of ϕ lying between the relative volatilities of all components:

$$\alpha_{\text{LNK},1\text{-ref}} > \phi_1 > \alpha_{\text{LNK},2\text{-ref}} > \phi_2 > \alpha_{\text{sandwich-ref}} > \phi_3 > \alpha_{\text{LK-ref}} > \phi_4 > \alpha_{\text{HK-ref}} > \phi_5 > \alpha_{\text{HNK},1\text{-ref}}$$

This gives $C - 1$ valid roots. Now write Eq. (7-25a) $C - 1$ times—once for each value of ϕ . We now have $C - 1$ equations and $C - 1$ unknowns (V_{min} and $Dx_{D,i}$ for all LNKs, sandwich components, and HNKs). Solve these simultaneous equations and then obtain D from Eq. (7-31) and L_{min} from Eq. (7-32). Problem 7.D15 is a sandwich component problem that must use this approach.

In general, Eq. (7-28) is of order C in ϕ where C is the number of components. Saturated liquid and saturated vapor feeds are special cases and, after simplification, are of order $C-1$. If the resulting equation is quadratic, the quadratic formula can be used to find the roots. Otherwise, a root-finding method or Goal Seek or Solver should be employed. If only one root, $\alpha_{\text{LK-ref}} > \phi > \alpha_{\text{HK-ref}}$, is desired, a good first guess is to assume $\phi = (\alpha_{\text{LK-ref}} + \alpha_{\text{HK-ref}})/2$. If looking for multiple roots, a good first guess to find the ϕ value between two α values is the average of the two α values.

The results of the Underwood equations are accurate only if the assumptions of constant relative volatility and CMO are valid. For small variations in α , a geometric average calculated as

$$\alpha_{i\text{-ref}} = (\alpha_{\text{bot}} \alpha_{\text{dist}})^{1/2} \text{ or } \alpha_{i\text{-ref}} = (\alpha_{\text{bot}} \alpha_{\text{feed}} \alpha_{\text{dist}})^{1/3} \quad (7-33)$$

can be used as an approximation.

EXAMPLE 7-2. Underwood equations

For the distillation problem given in Example 7-1, find the minimum reflux ratio. Use a basis of 100.0 kmol/h of feed.

Solution

- A. Define. The problem is sketched in Example 7-1. We now wish to find $(L/D)_{\min}$.
- B. Explore. Because the relative volatilities are approximately constant, the Underwood equations can be used to estimate the minimum reflux ratio.
- C. Plan. Because by most definitions in Problem 7D10 benzene is distributing in Example 7-1, Case A does not apply. Either Case B or C can be used. We follow a Case B analysis and use the $Dx_{D,i}$ values calculated in Example 7-1. We then solve Eq. (7-28) for ϕ value between the relative volatilities of the two keys $0.21 < \phi < 1.00$. Then V_{\min} can be found from Eq. (7-25a), D from Eq. (7-31), and L_{\min} from Eq. (7-32).
- D. Do it. Because the feed is a saturated vapor, $q = 0$, $\Delta V_{\text{feed}} = F(1 - q) = F = 100$, and Eq. (7-28) becomes

$$100 = \frac{2.25(40)}{2.25 - \phi} + \frac{1.0(30)}{1.0 - \phi} + \frac{0.21(30)}{0.21 - \phi}$$

Solving for ϕ between 0.21 and 1.00, we obtain $\phi = 0.5454$. Equation (7-25) is

$$V_{\min} = \sum_{i=1}^c \frac{\alpha_{i,\text{ref}} (Dx_{D,i})}{\alpha_{e\text{-ref}} - \phi} \text{ with } Dx_{D,i} = F z_i (FR_{D,i}).$$

For Case B analysis, the fractional recovery of benzene is the value calculated in Example 7-1 at total reflux, $Dx_{D,\text{ben}} = 100(0.4)(0.9985) = 39.94$. The other values are $Dx_{D,\text{tol}} = 100(0.3)(0.95) = 28.5$ and $Dx_{D,\text{cum}} = 100(0.3)(0.02) = 0.60$. Summing the three distillate flows, $D = 69.04$. Equation (7-25) becomes

$$V_{\min} = \frac{(2.25)(39.94)}{2.25 - 0.5454} + \frac{(1.0)(28.5)}{1.0 - 0.5454} + \frac{(0.21)(0.6)}{0.21 - 0.5454} = 115.04$$

From a mass balance, $L_{\min} = V_{\min} - D = 46.00$, and $(L/D)_{\min} = 0.6663$.

- E. Check. The Case A calculation gives essentially the same result.
- F. Generalize. The addition of more components does not make the calculation significantly more difficult as long as the fractional recoveries can be accurately estimated. The value of ϕ must be determined accurately because small errors can have a major effect on the results. Because this separation is easy, $(L/D)_{\min}$ is quite small, and $(L/D)_{\min}$ is not as dependent on the exact value of ϕ as it is when $(L/D)_{\min}$ is large.

7.3 GILLILAND CORRELATION FOR NUMBER OF STAGES AT FINITE REFLUX RATIOS

A general shortcut method for determining the number of stages required for a multicomponent distillation at finite reflux ratios would be extremely useful. Unfortunately, such a method has not been developed. However, Gilliland (1940) noted that he could empirically relate the number of stages N at finite reflux ratio L/D to the minimum number of stages N_{\min} and the minimum reflux ratio $(L/D)_{\min}$. Gilliland did a series of accurate stage-by-stage calculations and found that he could develop a graphical correlation of the function

$$Y = \frac{(N - N_{\min})}{(N + 1)} \quad (7-34a)$$

with the function

$$X = \frac{\left[\frac{L}{D} - \left(\frac{L}{D} \right)_{\min} \right]}{\left(\frac{L}{D} + 1 \right)} \quad (7-34b)$$

Since 1940, a number of investigators have developed equations to fit Gilliland's data so that the correlation can easily be used with calculators and computers (Coker, 2010; Davis, 2020a). The latest and possibly the best equation was developed by Davis (2020a, 2020b) using a rational function instead of a polynomial. Davis's correlation (2020b) is

$$Y = \frac{1 - X^{0.0031}}{1 - 0.99357X^{0.0031}} \quad (7-35)$$

Figure 7-3 shows Gilliland's data points and Davis's rational function. The data points are the result of Gilliland's stage-by-stage calculations and show the scatter inherent in this correlation.

Sometimes the values of N and N_{\min} , and thus Y , will be known and we will want to determine X and L/D . One advantage of Eq. (7-35) is that the inverse equation is easily determined:

$$X = \left(\frac{1 - Y}{1 - 0.99357Y} \right)^{1/0.0031} \quad (7-36)$$

Note that in Davis's paper (2020a), the value 0.99357 is truncated to 0.99. Equations (7-35) and (7-36) are extremely sensitive to this value, and the value 0.99 does not fit the curve.

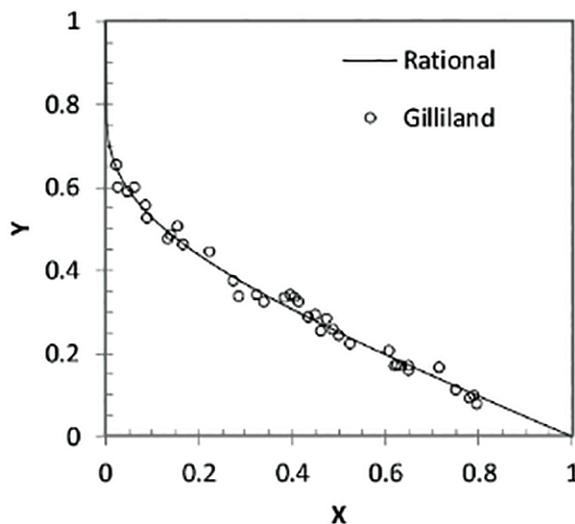


FIGURE 7-3. Gilliland correlation with equation (7-35) developed by Davis (2020a). Reprinted with permission from *Chemical Engineering Education*, 54(4), 219 (2020), copyright 2020, Chemical Engineering Education

Another advantage of Eq. (7-35) is that the function has the correct limiting behavior. As $X \rightarrow 0$, $Y \rightarrow 1$; and as $X \rightarrow 1$, $Y \rightarrow 0$ (see Problem 7A7).

To use the graphical Gilliland correlation or Davis's fit to the Gilliland correlation, we proceed as follows:

1. Calculate N_{\min} from the Fenske equation.
2. Calculate $(L/D)_{\min}$ from the Underwood equations or analytically for a binary system.
3. Choose actual (L/D) . This is usually done as some multiplier (1.05 to 1.5) times $(L/D)_{\min}$.
4. Calculate the abscissa X .
5. Determine the ordinate value Y .
6. Calculate the actual number of stages, N .

The Gilliland correlation should be used only for rough estimates. The calculated number of stages can be off by $\pm 30\%$, although they are usually within $\pm 7\%$. Because L/D is usually a multiple of $(L/D)_{\min}$, $L/D = M(L/D)_{\min}$, the abscissa can be written as

$$\text{abscissa} = X = \frac{[L/D - (L/D)_{\min}]}{L/D + 1} = \frac{M - 1}{[1 / (L/D)_{\min}] + M} \quad (7-37)$$

The abscissa is not very sensitive to the $(L/D)_{\min}$ value but does depend on the multiplier M .

The optimum feed plate location can also be estimated. First, use the Fenske equation to estimate where the feed stage would be at total reflux. This can be done by determining the number of stages required to go from the feed concentrations to the distillate concentrations for the keys.

$$N_{F,\min} = \frac{\ln \left[\frac{(x_{LK} / x_{HK})_{\text{dist}}}{(z_{LK} / z_{HK})} \right]}{\ln \alpha_{LK-HK}} \quad (7-38a)$$

Now assume that the relative feed location is constant as we change the reflux ratio from total reflux to a finite value. Thus,

$$\frac{N_{F,\min}}{N_{\min}} = \frac{N_F}{N} \quad (7-38b)$$

The actual feed stage can now be estimated from Eq. (7-38b). Because the estimate is based on an assumption that certainly may not be true and it does not include the effect of feed quality, it is not too accurate. Best practice is to use these estimates as first guesses of the feed location for simulations. Erbar and Maddox (1961; see King, 1980 or Coker, 2010) developed a somewhat more accurate correlation that uses more than one curve.

A rough heuristic is to estimate $N = 2.5 N_{\min}$. This estimate then requires *only* a calculation of N_{\min} and is useful for very preliminary estimates.

EXAMPLE 7-3. Gilliland correlation

Estimate the total number of equilibrium stages and the optimum feed plate location required for the distillation problem presented in Examples 7-1 and 7-2 if the actual reflux ratio is set at $L/D = 2$.

Solution

A. Define. The problem was sketched in Examples 7-1 and 7-2. $F = 100$, $L/D = 2$, and we wish to estimate N and N_F .

- B. Explore. An estimate can be obtained from the Gilliland correlation, while a more exact calculation could be done with a process simulator.
- C. Plan. Calculate the abscissa X from Eq. (7-34b), determine the ordinate Y from Davis's fit of the Gilliland correlation Eq. (7-35), and then find N from Eq. (7-34a). $(L/D)_{\min} = 0.6663$ was found in Example 7-2, and $N_{\min} = 3.77$ in Example 7-1. The feed plate location is estimated from Eqs. (7-38a) and (7-38b).
- D. Do it.

$$X = \frac{\frac{L}{D} - \left(\frac{L}{D}\right)_{\min}}{\frac{L}{D} + 1} = \frac{2 - 0.6663}{2 + 1} = 0.4446$$

The corresponding ordinate Y can be determined from Eq. (7-35):

$$Y = \frac{1 - X^{0.0031}}{1 - 0.99357X^{0.0031}} = \frac{1 - (0.4446)^{0.0031}}{1 - 0.99357(0.4446)^{0.0031}} = 0.2813$$

Solving for N in Eq. (7-34a) with Y known and $N_{\min} = 3.77$,

$$N = (Y + N_{\min}) / (1 - Y) = (0.2813 + 3.77) / (1 - 0.2813) = 5.64$$

Use $N = 6$.

$$N_{F,\min} \text{ is calculated as } N_{F,\min} = \frac{\ln \left[\frac{(x_{LK} / x_{HK})_{\text{dist}}}{(z_{LK} / z_{HK})} \right]}{\ln \alpha_{LK-HK}} = \frac{\ln \left[\frac{(0.413 / 0.0087)_{\text{dist}}}{(0.3 / 0.3)} \right]}{\ln(1.0 / 0.21)} = 2.49$$

where $x_{D,LK}$ and $x_{D,HK}$ were found in Example 7-1.

Then from Eq. (7-38b),

$$N_F = N \frac{N_{F,\min}}{N_{\min}} = 5.64 \left(\frac{2.49}{4.38} \right) = 3.20 \text{ use stage 3.}$$

- E. Check. A complete check requires solution with a process simulator.
- F. Generalize. The Gilliland correlation is a rapid method for estimating the number of equilibrium stages in a distillation column. It should not be used for final designs because of its inherent inaccuracy.

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PROBLEMS

A. Discussion Problems

- A1.** The Fenske equation:
- Is valid only for binary systems.
 - Was derived for minimum reflux.
 - Requires CMO.
 - Requires constant K values.
 - All of the above.
 - None of the above.
- A2.** If you want to use an average relative volatility, how do you calculate it for the Underwood equation?
- A3.** Develop your key relations chart for this chapter.
- A4.** In multicomponent distillation the Fenske equation can be used to:
- Estimate the fractional recoveries of the NKs at total reflux.
 - Calculate the number of equilibrium contacts at minimum reflux.
 - Estimate the average K value of the LK at total reflux.
 - All of the above.
 - None of the above.
- A5.** With the ready availability of process simulators, why do chemical engineers still use the Fenske-Underwood-Gilliland (FUG) method?
- A6.** Suppose you are doing a ternary distillation where component B, the LK, has a 98.3% recovery in the distillate, and component C, the HK, has a 99.8% recovery in the bottoms. If $\alpha_{A\text{-ref}} = \alpha_{B\text{-ref}}$, how does component A distribute?
- A7.** In Davis's fit for the Gilliland correlation, what are the values of N and L/D when $X \rightarrow 0$ and $Y \rightarrow 1$? What are the values of N and L/D when $X \rightarrow 1$ and $Y \rightarrow 0$?
- A8.** An engineer claims that fit A of the Gilliland correlation is better than fit B because when they compared the predictions of both fits to detailed simulator results for a separation of interest, fit A was closer than fit B. Have they proved that fit A is better? Explain your answer.

C. Derivations

- C1.** Derive a form of Eq. (7-13) for $(FR_{NK,bot})$ in terms of $(FR_{LK,dist})$.
- C2.** Explore the sensitivity of Eq. (7-35) in Example 7-3 at $X = 0.455$ by determining Y and N as the value of the constant 0.99357 changes. Try constant values of 0.990, 0.993, and 0.994.
- C3.** If the pinch point occurs at the feed point, mass balances can be used to find the minimum flows. Derive these equations. Note: A pinch point at the feed can occur but is unusual in multicomponent distillation.
- C4.** The choice of developing the Underwood equations in terms of V_{min} instead of solving for L_{min} is arbitrary. Rederive the Underwood equations solving for L_{min} and \bar{L}_{min} . Develop the equations analogous to Eqs. (7-25a) and (7-28).
- C5.** For binary systems, Eq. (7-28) simplifies to a linear equation for both saturated liquid and saturated vapor feeds. Prove this statement.
- C6.** If NKs do not distribute, you solve the Underwood Eq. (7-28) for ϕ , which satisfies $\alpha_{LK\text{-ref}} > \phi > \alpha_{HK\text{-ref}}$. However, if a different reference component is chosen for calculation of the relative volatilities, the value of ϕ changes. Despite the change in ϕ , V_{min} calculated from Eq. (7-25a) is unchanged. The proof that this is true is challenging for the general case but is tractable for a binary system with a saturated liquid feed because Eq. (7-28) becomes linear. Prove for a binary system with a saturated liquid feed that the solution for V_{min} is not affected by the choice of reference component for relative volatilities.

D. Calculation Problems

**Answers to problems with an asterisk are at the back of the book.*

- D1.** A distillation column separates 100.0 kmol/day of a saturated liquid feed that is 20.0 mol% ethanol (E), 35.0 mol% n-propanol (P), and 45.0 mol% n-butanol (But). Fractional recovery of butanol in bottoms = 0.972. Bottoms mole fraction butanol $x_{B,But} = 0.986$. Assume relative volatilities are constant: $\alpha_{E-But} = 4.883$, $\alpha_{P-But} = 2.336$, and $\alpha_{But-But} = 1.0$.
- Determine the flow rates of bottoms, B, and of distillate, D, in kmol/day; and determine the mole fractions of E, P, and But in the bottoms and in the distillate.
 - Find the minimum number of stages, N_{min} , required for this separation.
 - List any assumption(s) you have made and justify why they are reasonable. Note that the strongest justification is a calculation, not just words.
- D2.** We are separating a mixture of propylene, propane, and isobutane in a distillation column with a partial condenser and a partial reboiler at a pressure of 15.0 bar. We desire a 0.999 fractional recovery of propylene in the distillate, at least a 0.950 fractional recovery of propane in the bottoms, and at least a 0.9999 fractional recovery of isobutane in the bottoms. How many stages are required at total reflux?
Data: At 34°C, K Propylene = 1.00, K propane = 0.89, K isobutane = 0.42.
At 40°C, K propylene = 1.13, K propane = 1.000, and K isobutane = 0.46.
- D3.*** A special column acts as exactly three equilibrium stages. Operating at total reflux, we measure vapor composition leaving the top stage and the liquid composition leaving the bottom stage. The column is separating phenol from o-cresol. We measure a phenol liquid mole fraction leaving the bottom stage of 0.36 and a phenol vapor mole fraction leaving the top stage of 0.545. What is the relative volatility of phenol with respect to o-cresol?
- D4.** Separate 1,2-dichloroethane from 1,1,2-trichloroethane at 1 atm. Distillate is 99.15 mol% 1,2-dichloroethane, and bottoms is 1.773 mol% 1,2-dichloroethane. Saturated liquid feed is 60.0 mol% 1,2-dichloroethane. Relative volatility is approximately constant, $\alpha = 2.4$.
- Find the minimum number of stages using the Fenske equation.
 - Calculate L/D_{min} .
 - Estimate the actual number of stages for $L/D = 2.2286$ using the Gilliland correlation.
 - A detailed simulation gave 99.15 mol% 1,2-dichloroethane in the distillate, 1.773 mol% 1,2-dichloroethane in the bottoms for $L/D = 2.2286$, $N = 25$ equilibrium contacts, and optimum feed location is 16 equilibrium contacts from the top of the column. Compare this N with part c, and calculate the percent error in the Gilliland prediction.
- D5.** A distillation column will separate 100.0 kmol/h of a saturated liquid feed at 200 kPa that is 20.0 mol% propane (Pro), 35.0 mol% n-pentane (Pen), and 45.0 mol% n-hexane (Hex). The column has a total condenser and a partial reboiler. We want a fractional recovery of Hex in the bottoms = 0.983 and a fractional recovery of Pen in the distillate of 0.967.
- Make an appropriate assumption, and determine the flow rates of bottoms, B, and of distillate, D, in kmol/h; and determine the mole fractions of bottoms and of distillate.
 - Determine the bubble-point temperature of the feed, and calculate relative volatilities at this temperature. Use Pen as your reference component. Report the bubble-point temperature, the K values, and the values of relative volatilities. Use a DePriester chart or Eq. (2-28). Show your work.
 - Assume the relative volatilities found in part b are constant, and determine the minimum number of stages, N_{min} , required for this separation.
 - Do a calculation that justifies why the assumption made in part a is reasonable.
- D6.** A mixture of acetone and ethanol is distilled at 1.0 atm in a distillation column with a total condenser and a partial reboiler. We desire a distillate that is 0.999 mole fraction acetone and a bottoms that is 0.0013 mole fraction acetone. Feed is 40 mol % acetone,

it is a two-phase mixture that is 5/6 liquid, and feed flow rate is 50 mol/h. Data are in Problem 4.D7.

- a. Determine the relative volatility near the top of the column, near the bottom of the column, and near the intersection of the feed line and the equilibrium curve. Calculate the appropriate average relative volatility.
 - b. Use the Fenske equation to determine the number of equilibrium contacts at total reflux.
 - c. Assume CMO is valid and calculate the value of $(L/D)_{\min}$ from the McCabe-Thiele diagram.
 - d. Use the Gilliland correlation (or the Davis equation) to estimate the number of stages if $L/D = 1.05 (L/D)_{\min}$.
 - e. Estimate the optimum feed stage location.
- D7.** Your boss wants some idea of how expensive it will be to distill 155.0 kmol/h of a saturated liquid feed that is 5.0 mol% methane, 10.0 mol% ethane, 15.0 mol% n-butane, 22.0 mol% n-pentane, 22.0 mol% n-hexane, and 26.0 mol% n-heptane. Column pressure is 700.0 kPa. The column has a partial condenser and a partial reboiler. We want to recover 99.0% of the n-butane in the distillate and 98.3% of the n-pentane in the bottoms. Do the calculations of the K values either from the DePriester chart or from Eq. (2-28).
- a. Assuming that NKs do not distribute, calculate the values of D and B in kmol/h and the mole fractions in distillate and bottoms.
 - b. Do a bubble-point calculation at the feed conditions. Calculate the relative volatilities of all components with respect to the HK (n-pentane). Use these values as the average value of relative volatility for the entire column. Also determine the bubble-point temperature of the distillate to see if condensation will be expensive.
 - c. Determine the minimum number of stages for this separation with the Fenske equation.
 - d. Determine the minimum reflux ratio, $(L/D)_{\min}$, with the Underwood method.
 - e. Estimate the number of stages required if $L/D = M \times (L/D)_{\min}$ with the Gilliland correlation (Davis's fit is convenient) where $M = 1.04, 1.10, \text{ and } 2.0$.
 - f. Will this distillation be reasonably economical, or should an alternative be found? Briefly explain your reasoning.

Note: Parts b and d are easier to do with a spreadsheet or Wolfram.

- D8.*** We wish to separate a mixture of 40.0 mol% benzene and 60.0 mol% ethylene dichloride in a distillation column with a partial reboiler and a total condenser. The feed rate is 750 mol/h, and the feed is a saturated vapor. We desire a distillate product of 99.2 mol% benzene and a bottoms product that is 0.5 mol% benzene. Reflux is a saturated liquid, and CMO can be used. Equilibrium data can be approximated with an average relative volatility of 1.11 (benzene is more volatile).
- a. Find the minimum external reflux ratio.
 - b. Use the Fenske equation to find the number of stages required at total reflux.
 - c. Estimate the total number of stages required for this separation using the Gilliland correlation for $L/D = 1.2(L/D)_{\min}$.
- D9.** We are separating a mixture of benzene, toluene, and xylene in a distillation rectifying column. The column has a total condenser and no reboiler. The feed is a saturated vapor that is fed into the bottom stage of the column, flow rate $F = 150$ kmol/h, and feed is 52.0 mol % benzene, 38.5 mol % toluene, and remainder xylenes. Pressure is 1.0 atm, CMO is valid, and the relative volatilities are constant: $\alpha_{\text{Ben-Tol}} = 2.22, \alpha_{\text{Tol-Xy}} = 2.01$. The column is at 1.0 atm. The reflux ratio $L/D = 9$, and the distillate is 0.007 mole fraction toluene.
- a. Based on the best assumption you can make, use mass balances and CMO to calculate: B, mole fractions in bottoms, D, and mole fractions in distillate.

- b. Although the column has a feed and bottoms removal, we can still operate at total reflux ($D = 0$ so that $L/V = 1$). At total reflux, how many stages are required to obtain the separation achieved in part a?
- c. Use the Fenske equation to estimate xylene mole fraction in the distillate.
- d. What is the minimum reflux ratio for separation in part a, but with xylene distillate mole fraction from part c?
- e. Use the Gilliland correlation to estimate the actual number of stages if $L/D = 9$.
- D10.** When is a non-key distributing, and when is it nondistributing? For almost all chemicals, five 9s purity {concentrations of impurity below 10.0 ppm, mass (mass fraction $< 1.0 \times 10^{-5}$) or [approximately mole fraction $< (1.0 \times 10^{-5})$]} would be low enough that the chemical can be accepted in the product and could be considered to be nondistributing (this is a very tight definition of nondistributing). A less strict concentration limit (four 9s purity) would use 100.0 ppm. Another possible definition of nondistributing that is less strict is a concentration that causes less than a 0.1% or 0.01% change in calculations of other variables (e.g., concentrations of other components, flow rates, equilibrium behavior, and so forth). Return to Example 7-1 and use the Fenske equation to explore under what conditions the NK benzene can be considered nondistributing.
- a. Redo the calculations in Example 7-1 by assuming benzene is nondistributing, and determine new flow rates of D and B and new mole fractions in distillate and bottoms. Are the percentage changes in flow rates of B and D and percentage change in mole fractions of toluene and cumene small enough that benzene could be considered nondistributing by either the 0.1% or the 0.01% criteria?
- b. Redo the calculations in Example 7-1 with $FR_{\text{tol,dist}} = FR_{\text{cum,bot}}$. Find the value of $FR_{\text{tol,dist}} = FR_{\text{cum,bot}}$ and the corresponding value of N_{min} at which benzene first meets the 10.0 ppm criterion for nondistributing.
- c. Redo the calculations in Example 7-1 with $FR_{\text{tol,dist}} = FR_{\text{cum,bot}}$. Find the value of $FR_{\text{tol,dist}} = FR_{\text{cum,bot}}$ and the corresponding value of N_{min} at which benzene first meets the change of less than 0.01% in toluene and cumene distillate and bottoms concentrations criterion for nondistributing.
- d. Find the value of $FR_{\text{tol,dist}} = FR_{\text{cum,bot}}$ and the corresponding value of N_{min} at which benzene first meets the change of less than 0.1% in toluene and cumene distillate and bottoms concentrations criterion for nondistributing.
- D11.** We simulate a distillation column and find we can obtain the desired separation with 31 stages plus a partial reboiler and a total condenser if we use an $L/D = 3$. With total reflux, we find that the desired separation is obtained with 13 stages plus a partial reboiler and a total condenser. Estimate $(L/D)_{\text{min}}$.
- D12.** A distillation column is separating toluene and xylene, $\alpha = 3.03$. The feed is a saturated liquid, and reflux is returned as a saturated liquid. $p = 1.0$ atm. $F = 100.0$ kmol/h. Distillate mole fraction is $x_D = 0.996$, and bottoms $x_B = 0.008$. Use the Underwood equation to find $(L/D)_{\text{min}}$ and V_{min} at feed mole fractions of $z = 0.1, 0.3, 0.5, 0.7$, and 0.9 . Check your result at $z = 0.5$ with a McCabe-Thiele diagram. What are the trends for $|Q_{c,\text{min}}|$ and $Q_{R,\text{min}}$ as the toluene feed concentration increases? Hint: If you write the Underwood equation and solve *algebraically* for ϕ , the problem is easier than it looks.
- D13.*** We have a column separating benzene, toluene, and cumene. The column has a total condenser, a total reboiler, and nine equilibrium stages. The feed is 25.0 mol% benzene, 30.0 mol% toluene, and 45.0 mol% cumene. Feed rate is 100 mol/h, and the feed is a saturated liquid at 1.0 atm. The column pressure is 1.0 atm. The equilibrium data can be represented as constant relative volatilities: $\alpha_{\text{BT}} = 2.5$, $\alpha_{\text{TT}} = 1.0$, and $\alpha_{\text{CT}} = 0.21$. We desire 99.0% recovery of toluene in the distillate and 98.0% recovery of cumene in the bottoms.
- a. Determine the required external reflux ratio.
- b. If $\alpha_{\text{BT}} = 2.25$ instead of 2.5, what is the value of L/D ?

- D14.** A distillation column is separating 100.0 kmol/h of a saturated vapor feed that is 30.0 mol% ethanol, 25.0 mol% i-propanol, 35.0 mol% n-propanol, and 10.0 mol% n-butanol at a pressure of 1.0 atm. We want a 98.6% recovery of i-propanol in the distillate and 99.2% recovery of n-propanol in the bottoms. The column has a total condenser and a partial reboiler. For parts b, c, and d, use the FUG method. If we choose n-propanol as the reference, the relative volatilities are ethanol = 2.17, i-propanol = 1.86, n-propanol = 1.0, and n-butanol = 0.412. These relative volatilities can be assumed to be constant.
- Find D , B , $x_{D,i}$, and $x_{B,i}$.
 - Find N_{\min} and $N_{F,\min}$.
 - Find $(L/D)_{\min}$. A spreadsheet is highly recommended to find ϕ .
 - If $L/D = 1.10(L/D)_{\min}$, find N and the feed stage.
- D15.*** A distillation column is separating benzene ($\alpha = 2.25$), toluene ($\alpha = 1.00$), and cumene ($\alpha = 0.21$). The column is operating at 101.3 kPa. The column has a total condenser and a partial reboiler, and the optimum feed stage is used. Reflux is a saturated liquid, and $L_0/D = 1.2$. Feed rate is 1000.0 kmol/h. The saturated liquid feed is 39.7 mol% benzene, 16.7 mol% toluene, and 43.6 mol% cumene. Recover 99.92% of the benzene in the distillate and 99.99% of the cumene in the bottoms. For a first guess to this design problem, use the FUG approach to estimate the optimum feed stage and the total number of equilibrium stages. Note: The Underwood equations must be treated as a Case C problem.
- D16.*** We are separating a mixture of ethanol and n-propanol. Ethanol is more volatile, and the relative volatility is approximately constant at 2.10. The feed flow rate is 1000.0 kmol/h. The feed is 60 mol% ethanol and is a saturated vapor. We desire $x_D = 0.99$ mole fraction ethanol, and $x_B = 0.008$ mole fraction ethanol. The reflux is a saturated liquid. There are 30 stages in the column (including the partial reboiler). Use the FUG approach to determine:
- The number of stages (including partial reboiler) at total reflux.
 - $(L/D)_{\min}$.
 - $(L/D)_{\text{actual}}$.
- D17.** A distillation column operating at 200 kPa separates 100 kmol/h of a saturated liquid feed at 200 kPa that is 20 mol% propane (Pro), 35 mol% n-pentane (Pen), and 45 mol% n-hexane (Hex). The column has a total condenser and a partial reboiler. We want a fractional recovery of Hex in the bottoms = 0.983, and a fractional recovery of Pen in the distillate of 0.967.
- Make an appropriate assumption and determine flow rates of bottoms, B , and of distillate, D , in kmol/h; and determine mole fractions of bottoms and of distillate.
 - Determine bubble-point temperature of feed and calculate relative volatilities at this temperature. Use Pen as your reference component. Report bubble-point temperature, K values, and values of relative volatilities. Use DePriester charts or Eq. (2-28). Show your work.
 - Assume relative volatilities found in part b are constant, and determine minimum number of stages, N_{\min} , required for this separation.
 - Do a calculation that shows assumption made in part a is correct.
- D18.** A depropanizer has the following feed and constant relative volatilities:
- | | |
|---------------|---------------------------------------|
| Methane (M): | $z_M = 0.229$, $\alpha_{M-P} = 9.92$ |
| Propane (P): | $z_P = 0.368$, $\alpha_{P-P} = 1.00$ |
| n-Butane (B): | $z_B = 0.322$, $\alpha_{B-P} = 0.49$ |
| n-Hexane (H): | $z_H = 0.081$, $\alpha_{H-P} = 0.10$ |
- Reflux is a saturated liquid. The feed is a saturated liquid fed at 1.0 kmol/(unit time). Assume CMO.
- $L/D = 1.5$, $FR_{P,\text{dist}} = 0.9854$, $FR_{B,\text{bot}} = 0.8791$. Use the FUG method to estimate N .
 - If $N = 20$, $FR_{P,\text{dist}} = 0.9854$, and $FR_{B,\text{bot}} = 0.8791$, estimate the required L/D .

- c. Find the split of normal hexane at total reflux using N_{\min} .
- d. $L/D = 1.5$, $FR_{p,\text{dist}} = 0.999$, $FR_{B,\text{bot}} = 0.8791$. Use the FUG method to estimate N .
- Note: Do part a first. Parts of the solution of part a can be reused for the other parts b to d.
- D19.** A distillation column with a partial reboiler and a total condenser operating at 7.0 bar is separating 100.0 kmol/h of a saturated liquid feed that is 25.0 mol% ethane (C2), 35.0 mol% n-butane (C4), and 40.0 mol% n-pentane (C5). CMO can be assumed valid, and assume that ethane does not distribute. We want 99.2% recovery of n-butane in the distillate and 98.3% recovery of n-pentane in the bottoms. The K values at the distillate are $K_{C2} = 5.56$, $K_{C4} = 0.655$, and $K_{C5} = 0.234$. The K values at the bottoms are $K_{C2} = 10.67$, $K_{C4} = 2.21$, and $K_{C5} = 0.993$. Use the correct average for relative volatilities based on the values of the relative volatilities calculated at distillate and bottoms.
- a. Find the distillate mole fractions and the value of the distillate flow rate.
- b. Find N_{\min} .
- c. Find $(L/D)_{\min}$.
- d. Find N from the Gilliland correlation for $M = 1.2$.

E. More Complex Problems

- E1.** We are separating 100.0 kmol/h of a saturated liquid feed that is 45.0 mol% propane (P), 15.0 mol% n-butane (B), and 40.0 mol% n-hexane (H). Relative volatilities are $\alpha_{P-P} = 1.0$, $\alpha_{B-P} = 0.49$, $\alpha_{H-P} = 0.10$. At minimum reflux, we want fractional recoveries $FR_{P,\text{dist}} = 0.995$ and $FR_{H,\text{bot}} = 0.998$. Find $(L/D)_{\min}$, flow rates D and B , and mole fractions of three components in distillate and bottoms at minimum reflux.
- E2.** Separation of propylene from propane is a very important but expensive distillation. Your boss wants to know the effect of changing the column pressure on the number of stages and on the temperature at which the distillate condenses. Use the FUG method to estimate the number of stages required for the following feeds listed. In all cases, we want 99.8 mol% propylene in the distillate with a 99.1% recovery of propylene in the distillate. Use either DePriester charts or Eq. (2-28) for K values. Operate with $L/D = 1.05(L/D)_{\min}$. Report the temperature of the distillate, the minimum number of stages, $(L/D)_{\min}$, and predicted N_{actual} . Calculate the average relative volatility,

$$\bar{\alpha} = \sqrt{\alpha(T_{\text{distillate}}) \times \alpha(T_{\text{bottoms}})} \quad \text{where } \alpha = K_{\text{propylene}} / K_{\text{propane}}$$

- a. Saturated liquid feed, 50 mol% propylene, column pressure is 22.0 bar.
- b. Saturated liquid feed, 50 mol% propylene, column pressure is 7.0 bar.
- c. Saturated liquid feed, 50 mol% propylene, column pressure is 1.013 bar.
- d. Saturated vapor feed, 50 mol% propylene, column pressure is 7.0 bar.

Note: This problem can be solved by brute force, or it can be simplified first and then be easily solved.

F. Problems Requiring Other Resources

- F1.** What variables does the Gilliland correlation not include? How might some of these be included? Check the Erbar-Maddox (1961) method (see King, 1980, or Coker, 2010).

G. Computer Simulation Problems

- G1.** Repeat Problem 7.D12 on Aspen Plus using RadFrac and the Peng-Robinson correlation.
- a. Find N at total reflux (operate with very small feed and distillate rates and a large L/D).
- b. Find $(L/D)_{\min}$ accurately by simulating the process with a few hundred stages.
- c. Find the actual number of stages and the optimum feed stage at $L/D = 1.25(L/D)_{\min}$.
- G2.** Repeat Problem 7.G1 except using DSTWU in Aspen Plus (Lab 6) instead of RadFrac.

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