VOLUME 2 | FOURTH EDITION

Ludwig's Applied Process Design for Chemical and Petrochemical Plants

- Distillation • Petroleum fractionation
- Gas processing • Dehydration • Packed towers

A. Kayode Coker

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Dedication

He who bears within himself the firm volition for what is good, and strives to give purity to his thoughts has already found the way to the Highest. All else will then be added unto Him.

Keep the hearth of your thoughts pure, by so doing you will bring peace and be happy.

"In the Light of Truth" by Abd-ru-shin

In memory of Ernest E. Ludwig

(A great chemical engineer)

and

In loving memory of my parents

Mr. Gabriel Shodipo Coker and Mrs. Modupe Ajibike Coker

To my wife, Victoria

Love and thanks

Front Matter

Ludwig's Applied Process Design for Chemical and Petrochemical Plants

Volume 2. Fourth Edition

A. Kayode Coker

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Preface to the Fourth Edition

This complete revision of *Applied Process Design for Chemical and Petrochemical Plants, Volume II*, builds upon the late Ernest E. Ludwig's classic text to enhance its use as a chemical engineering process design manual of methods and proven fundamentals with supplemental mechanical and related data, nomographs and charts (some in the expanded appendices). Significantly expanded and thoroughly updated, this fourth edition contains new topics that will assist the engineer in examining and analyzing problems and finding design methods and mechanical specifications to secure the proper mechanical hardware to accomplish a process objective.

This latest edition includes improved techniques and fundamental design methodologies to guide the engineer in designing process equipment, such as distillation and absorption columns, and applying chemical processes to the properly detailed hardware (equipment). Like its predecessor, this edition continues to present updated information for achieving optimum operational and process conditions and avoiding problems caused by inadequate sizing and lack of internally detailed hardware. The various derived and proven equations have been employed in actual plant equipment design, process control and operator's training, and some of the most reasonable available to inexperienced and experienced engineers.

This book further provides both fundamental theories where applicable and directs application of these theories to applied equations essential in the design effort. This approach at presentation of design information serves well for troubleshooting distillation and absorption equipment/system performance analysis.

Chapter 10 on Distillation incorporates further insights in the theory

of chemical engineering thermodynamics, azeotropic mixtures, and simulation of thermodynamic programs. It includes equilibrium flash calculations, Smoker's equations for determining the number of theoretical plates at constant relative volatility, enlarged batch separation fundamentals, the Ponchon–Savarit method of binary mixture with expanded fundamental theory, and additional multicomponent systems. It provides details to distillation operation involving side streams; expanding on the sizing of towers with valve trays, detailing steps in achieving optimum number of trays; troubleshooting predictive maintenance and controls of distillation columns, further illustrations of Gamma scanning and heat integration in distillation columns employing the grand composite curve, distillation sequencing and capital cost considerations of distillation columns.

Chapter 11 is a new chapter that covers fractionation of crude oil, gas processing, dehydration technique, hydrocarbon absorption and sour water stripping. This chapter provides case studies using Honeywell design simulation software (UniSim™) to perform simulation of these processes. It includes a developed computer program for designing dehydrating equipment and a glossary.

Chapter 12 is a new chapter that features enhanced distillation techniques, such as pressure swing distillation, vapor recompression, extractive distillation, reactive distillation and residue curve maps. In reactive distillation, thermodynamic and reaction mechanism involving the production of methyl acetate are provided. Further, the chapter includes a case study using Honeywell UniSim™ design simulation software to simulate the reaction in the production of methyl acetate at specified conditions.

Chapter 13 deals with mechanical hardware and internal specifications of distillation columns, and provides charts and

specification forms that can be used to specify details required for manufacturing the hardware. A detailed distillation equipment diagram of a debutanizer unit with its specification can be accessed on the Elsevier website (www.elsevier.com).

Chapter 14 extensively reviews packed columns with various generalized pressured drop charts (GPDC), and provides detailed equations for sizing a packed column with random packings. The chapter includes a developed computer program for specifying the size of a column with different packing types. The chapter also provides new GPDCs for structured type packings, and simulation exercises using Honeywell UniSim™ software.

Software/Programs/Excel Spreadsheets/Charts

New additions to the fourth edition are Excel™ spreadsheet programs as worked examples, and developed computer programs (Absoft Fortran™) that use the Microsoft Runtime Windows Environment (MRWE) in various sections of the chapters. These programs are in executable format and Appendix I provides an illustration of its use.

Incorporated is Honeywell UniSim™ simulation design software to perform case studies and some worked examples in the text. The programs were developed using UniSim R360.1 and UniSim R370 build 13058 version and the text includes the solutions of these examples and case studies. Also, many vapor-liquid equilibrium (VLE) diagrams of binary mixtures have been developed using established literature data and Microsoft Excel spreadsheet program (eighty six figures). These diagrams are in appendix M and are readily available from the website.

A program on Conversion Table developed by Mr. Ahmed Mutawa from SASREF Co., is available. All of the above can be accessed from the Elsevier website (www.elsevier.com).

This volume includes new appendices J, K, L and M, new charts and updated references with corresponding websites for review. Appendix J contains figures of equilibrium K-values, appendix K reviews the description of Honeywell UniSim simulation software, appendix L contains simulation results of chapter 11 using UniSim simulation software and appendix M contains eighty six diagrams of vapor-liquid equilibrium (VLE) of binary mixtures.

The author assumes that the reader is an advanced undergraduate/graduate or equivalent chemical/process or related engineer, having a sound knowledge of the fundamentals of the profession. With this assumption, the author illustrates the techniques of design and mechanical details necessary for construction of processes. The aim of the process engineer is to ensure that results of his or her process calculations for equipment are specified in terms of something that can be economically constructed or selected from special designs or manufacturers. This edition follows the format of previous editions, and the concept is stressed to a reasonable degree in the various chapters.

The techniques of applied chemical plant process design continue to improve as the science of chemical engineering develops new and better interpretations of the fundamentals of chemistry, physics, metallurgy, mechanical engineering, and polymer/plastic science. Accordingly, this fourth edition presents additional reliable design methods based on sound experimental data, proven techniques developed by companies and individuals and groups considered competent in their subjects and who are supported by pertinent data. In many chemical and petrochemical processes, the designer will find design techniques adaptable to 75 to 80 percent of his/her requirements. Thus, an effort has been made to place this book in a position of utilization somewhere between a handbook and an applied

teaching text. The present work is considered suitable to provide a practical guide to chemical process design for advanced undergraduate and graduate students in chemical engineering, practicing process engineers, and chemical/process engineers working in process development. The text can readily be used, if a general course in plant design is available to fill in the broader factors associated with overall plant layout and planning.

To access additional material accompanying this book, please visit: http://www.elsevierdirect.com/companions/9780750683661

On this companion website there are many useful Excel spreadsheets, additional appendices, examples and software.

Acknowledgments

This project is a culmination of three years of research, collating relevant and recent materials from organizations, institutions, companies and publishers. Soliciting for proofreaders has also been an important step in the editorial process. Emulating the work of the late E. Ludwig is a formidable task, and without the help of various individuals and organizations, this project would have been impossible to achieve.

I would like to express my deepest gratitude to Dr. Clive J. Mumford for volunteering, once again to review chapter 10. I am grateful for his analyses, criticisms and suggestions. His views on various aspects in the chapter and his attention to intricate details on some diagrams are greatly appreciated. Sincere thanks also to Dr. James Ryan for his analyses and comments and invaluable suggestions. His willingness in helping to review the chapter and provide succinct comments and guidance has indeed helped in enhancing the contents of this edition.

Sincere appreciation and thanks are given to Mr. Amjed El-Zubeir of Honeywell Process Solutions and also to Honeywell Process Solutions Amjed, thanks for your patience and help in reviewing chapter 11 and providing guidance on the simulation exercises. Sincere gratitude goes to Honeywell Process Solutions for granting permission to incorporate the use of UniSim design simulation software in the book. Dr. P. A. Bryant, a professor at Louisiana State University, contributed to updating the Absorption and Stripping section in this chapter.

I would like to record my appreciation and thanks to Lee Partin and Bob Huss of Eastman Co., USA for providing useful comments on chapter 12 and my sincere gratitude also goes to Professor James Fair

of the University of Texas at Austin, USA for his expert opinions, comments and invaluable suggestions.

Mr. Joseph Rivera has provided some astonishing diagrams, seeking attention to details, and I am grateful for his professionalism. Thank you to Mr. Ahmed Mutawa for developing the Conversion Table program for this book.

Many organizations, institutions and companies such as Armfield Co. U.K., Gas Processors Suppliers Association (GPSA) USA, Eastman Co. USA, Honeywell Process Solutions, Sulzer ChemTech, Ltd. Switzerland, SASREF Co. Saudi Arabia, Absoft Corporation© USA, American Institute of Chemical Engineers, *Chemical Engineering magazine* by Access Intelligence, USA, and *Hydrocarbon Processing* have readily given permission for the use of materials, and their release for publication. I greatly acknowledge and express my deepest gratitude to these organizations.

I have been privileged to have met with nice people as Tim Calk and Phil Carmical, former editors of Gulf Publishing Company and Elsevier respectively. Phil first suggested Ludwig's classic work several years ago and defended my proposals in the presence of Elsevier's management to revise the three volumes of Ludwig. It's a formidable task, and thanks for your trust and encouragement, and to you Tim, I'm deeply grateful for your assistance since our work together on my second book project several years ago, and for your encouragement and assistance since.

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Finally,

Without Him, I am nothing

Life is in God alone

In Him alone is the energy that lies in life.

A. Kayode Coker

BIOGRAPHY

A. Kayode Coker, Ph.D., Formerly, Chairman of Chemical & Process Engineering Department at Jubail Industrial College, Saudi Arabia, is a consultant for AKC Technology in England. He is a Chartered Chemical Engineer, a Chartered Scientist and a Fellow of the Institution of Chemical Engineers, U.K., (C.Eng, CSci, FIChemE). He is also a Senior member of the American Institute of Chemical Engineers (AIChE). He holds a B.Sc. honors degree in Chemical Engineering, a Master of Science degree in Process Analysis and Development, and a Ph.D. degree in Chemical Engineering, all from Aston University, Birmingham, U.K., and a Teachers' Certificate in Education at the University of London, U.K. He has directed and conducted short courses in both the U.K. and for SABIC industries in Saudi Arabia. His articles have been published in several international journals, and an author of three books in Chemical Engineering and a contributor to the Encyclopaedia of Chemical Processing and Design, Vol. 61. He was named as one of the International Biographical Centre (Headquarters in Cambridge, U.K.) LEADING ENGINEERS of the World, 2008.

Chapter 10. Distillation

Part 1: Distillation Process Performance

An important unit operation in the chemical process industries (CPI) is to separate a mixture into its components. This is termed diffusional or mass transfer. A typical chemical plant, as illustrated in the schematic flow sheet of Figure 10-1 will consist of both reaction and separation units. The raw materials are first purified in a separating unit and then fed to the reactor. A factor representing the efficiency with which raw materials are converted to products is the selectivity. This is given by moles of primary product produced divided by the moles of limiting reactant consumed. It can vary between 0 to 1, depending on the stoichiometry, the molar feed ratio of reactants, reactor temperature(s), reactor configuration, the catalyst if required *etc*. Any unreacted feed that remains is separated from the reaction products and recycled back to the reactor. The products are further separated and purified, before being marketed or used in subsequent processes. If the products contain “noncondensable” components, such as methane, hydrogen, argon, these must be separated by flash separation or similar process. Before flash separation, the process stream is usually cooled and depressurized. Afterwards, it may be fed to another separator to remove and purify useful components. Any remaining raw materials are recycled to the chemical process. The arrangement of Figure 10-1 is typical of many chemical processes, and is illustrated in Hydrocarbon processing magazine [257] and many textbooks [258-261].

Figure 10-1

Schematic flow diagram of a chemical process.

The main techniques used for separations are distillation, absorption, liquid-liquid extraction, drying, leaching, crystallization and gas adsorption. These processes account for 40% to 70% of both the capital and operating costs in the CPI since they significantly affect energy consumption, product costs and manufacturing profits. About 43% of the energy consumed by the U.S. chemical process industries is used for separation processes [244].

The optimum method for a given process depends on the physical state (e.g. liquid, solid or gaseous) of the components such factors as the achievable degree of separation, the required volatility and toxicity of the components and purity criteria. In general, the choice of the key design parameters will need to balance costs incurred due to the raw materials, the reactor system and the separation processes. For example, in some cases, selectivity increases as conversion of the limiting reactant decreases. Increased selectivity means that less raw material is converted to byproduct, thereby saving on raw material costs. However, as conversion decreases, more unreacted raw material must be separated from the product stream before being recycled to the reactor. The related costs increase quite steeply as conversion decreases. Conversely, reactor volume and its cost increases very

sharply as conversion approaches unity. An optimum value therefore must balance raw material costs, reactor costs and separation costs. The chosen value will determine the stream temperature, pressure and composition as well as the size of both reactor and the separation system [262].

Although the art and science of distillation have been practiced for many years, studies continue to determine the best design procedures for multicomponent, azeotropic, batch, multidraw, multifeed and other designing. Some shortcut procedures are adequate for complex situations many systems, but have limitations in others.

The methods outlined in this chapter are considered adequate for the stated conditions, but there may be exceptions to these generalizations. The process engineer often “double checks” his or her results by using a second method to verify the “ball-park” results, or to check the findings of shortcut methods that lack fine detail.

Current design techniques using computer programs allow excellent prediction of performance for complicated multicomponent systems, such as azeotropic or high hydrocarbon content process streams as well as those which specify extremely high purity of one or more product streams. Of course, the more straightforward systems can be predicted with excellent accuracy. Such techniques allow examination of an array of variables; invaluable in selecting optimum or preferred conditions of operation.

The high manufacture costs associated with and installation of this equipment warrant the best design and checking techniques. The process diagram of Figure 10-2 will be used as reference for the systems and methods presented in this chapter Figure 10-2a shows fractionating columns in a chemical plant.

Figure 10-2

Schematic distillation tower/column arrangement with total condenser.

Figure 10-2a

A photograph of fractionating columns showing Demethanizer, Deethanizer, Depropanizer and Debutanizer units.

(courtesy of Sulzer Chemtecth Ltd.)

The right size of equipment should be chosen for the process. In addition, a choice between a stagewise or continuous, i.e., tray or packed process needs to be made. Trays act as individual stages and produce stepwise changes in concentration, whereas packed towers produce continous differential changes in concentration.

This chapter reviews the vapor-liquid equilibria of some representative mixtures, conditions of bubble of liquid mixtures and dew points of gaseous mixtures, isothermal equilibrium flash calculations, the design of distillation towers with valve trays, packed tower design, Smoker's equation for estimating the number of plates in a binary mixture, petroleum fractionation, gas processing, dehydration, reactive distillation, the computation of multicomponent recovery and minimum trays in distillation columns troubleshooting, distillation sequencing and heat integration of distillation columns. Computer programs and Honeywell UniSim simulation software are used to illustrate various examples in this volume.

10.1. Equilibria Basic Considerations

Distillation design is based on the theoretical premise that heat and mass transfer from stage to stage (theoretical) are in equilibrium [225-229]. Actual columns with actual trays are designed by establishing column tray efficiencies, and applying these to the theoretical trays or stages that have been determined by the calculation methods to be presented in later sections.

Dechman [109] illustrates a modification to the usual McCabeThiele diagram that allows for unequal molal overflow. Ryan [272] provides

an alternative McCabeThiele diagram for relative volatilities below 1.25. The procedure of generating the revised McCabeThiele diagram is provided later in this volume.

Distillation, extractive distillation, liquid-liquid extraction and absorption are all techniques used to separate binary and multicomponent mixtures of liquids and vapors. Reference 121 examines approaches to determine optimum process sequences for separating components from a mixture, primarily by distillation. The most important assumption is that vapor and liquid are in equilibrium at each theoretical stage. Heat and mass transfer are complete and at equilibrium, meaning that the temperature of both phases are the same. The composition in each phase is uniform.

Vapor-liquid equilibrium data is therefore essential to adequately perform distillation calculations and so to correlate composition, temperature, and system pressure.

There are two types of systems: ideal and nonideal. These terms apply to the simpler binary or two component systems as well as to the often more complex multicomponent systems.

Figure 10-3 shows a photograph of a typical vapor-liquid equilibrium apparatus with controls. Figure 10-4a and Figure 10-4b show a photograph of a pilot plant distillation column and its schematics respectively; Figure 10-4c and Figure 10-4d show a photograph of the column plates and its schematics with the direction of vapor and liquid flows.

Figure 10-3

A photograph of Vapor-Liquid Equilibrium apparatus

(courtesy of Saudi Shell-Aramco Refinery Co. (SASREF)).

Figure 10-4a

Photograph of a pilot plant distillation column

(courtesy of Armfield, U.K.)

Figure 10-4b

Schematics of a pilot plant distillation column

(courtesy of Armfield, UK).

Figure 10-4c

Photograph of the column plates

(courtesy of Armfield, U.K).

Figure 10-4d

Schematics of a pilot plant distillation column plates showing liquid and vapor movements

(courtesy of Armfield, UK). D = A plate; E = A Support rod; F = A weir; Liquid Movement ⬇; Vapour Movement ⬆; G = A downcomer; H = U-tube.

Figure 10-5 illustrates a typical volatility vapor-liquid equilibrium curve for a particular component in a distillation separation. Usually the more volatile component (MVC) of the mixture is the one of interest.

Figure 10-5

Continuous fractionation of binary mixtures; McCabeThiele diagram with total condenser.

10.2. Vapor-Liquid Equilibria

Vapor-liquid equilibria (VLE) data form the basis for evaluation of distillation operations; the most useful form of this data are boiling point diagrams. An ideal liquid or gas solution is one for which the activity coefficient is unity. There are few liquid systems that behave ideally at any pressure, and the assumption of ideal liquid behavior is generally correct only for members of homologous series that have similar molecular weights. At low pressure, the vapor phase of a mixture approaches ideal behavior and follows the ideal gas law. For these conditions, the fugacity is equated with the partial pressure and the standard state fugacity of the liquid becomes equal to the vapor pressure.

Figure 10-6a is a phase diagram for a mixture of methanol and water at a constant pressure of one atmosphere (101.3 kPa). The boiling points of pure methanol, the more volatile component (MVC) and pure water are 64.5°C and 100°C respectively. The lower curve represents the locus of the bubble point temperatures for all mixture of methanol and water. This curve is termed the saturated liquid (bubble point liquid); states of subcooled liquid lie below it.

Figure 10-6a

Phase diagram for the mixture of methanol and water at 1 atm.

The upper curve represents the locus of the dewpoint temperatures; states of superheated vapor lie above it. This curve is termed the saturated vapor (dewpoint vapor). Both of these curves originate at the boiling temperature of one constituent (100°C) and terminate at the boiling temperature of the other constituent (64.5°C). The twophase region lies between these curves.

Assume that a mixture of methanol-water of the composition of point *z* (50 mole %) exists at a temperature of 70°C (represented by point *f* on Figure 10-6a). As this mixture is heated, it will remain liquid until

point *d* (74°C) is reached, at which the first bubble of vapor is formed. This is the bubble-point temperature for a 50 mole % mixture at the system pressure of 1 atmosphere. The vapor phase composition at the bubble point is 78 mole % of methanol (point *e*). If heating of the mixture is continued, and the vapor generated is maintained in equilibrium with the liquid, the mixture reaches point *c* on the diagram, at which both vapor and liquid exist. The overall composition of the total mixture remains *z*, but the entire liquid portion will have a composition *x* (28 mole %), and the vapor portion will have a composition *y* (66 mole %). However, once a small amount of liquid has been vaporized, the remainder no longer contains 50 mole % methanol; it contains less, as the vapor produced has a higher concentration of this component. Consequently, the temperature of the system steadily rises as more and more liquid is vaporized, and the compositions of both phases change continuously during the process. The vapor and liquid phase compositions change as indicated by paths *bh* and *ag*, until the dew point is reached at point *g*, at which the last bubble of liquid disappears.

Correspondingly, consider an experiment started with 50 mole % mixture at 90°C (point *i*), where the mixture is 100% vapor. As the temperature is lowered, the dewpoint temperature is encountered at 85°C at system pressure (point *g*), and the first drop of liquid is formed – which is about 15 mole % (point *h*). As the temperature of the system steadily decreases, more and more vapor is condensed, and the compositions of both phases change continuously as indicated by paths *ga* and *hb*, until the bubble point is reached at point *d*, where the last bubbles of vapor disappear.

The bubble and dew temperatures are composition-dependent. For example, the bubble temperature of a 28 mole % mixture is 77°C (point *b*) and the bubble temperature of a 10 mole % is 87°C (point *j*). A similar discussion to that given above could be presented for the

dew temperatures of these systems. The bubble and dewpoint discussions could also be presented on the pressure diagram at constant temperature as shown in Figure 10-6b; however the references are to the bubble and dew pressures. The horizontal dotted lines connecting coexisting compositions are tie lines, and the relative amounts of the vapor and liquid present can be determined by the material balances.

Figure 10-6b

Illustration of a P-x-y diagram.

In Figure 10-6a, any point above the saturated vapor curve represents an all vapor mixture (superheated). Correspondingly, all points below the saturated liquid curve represent an all-liquid condition (subcooled), and points between the two curves correspond to a liquid-vapor mixture. Where the boiling points diagram is unavailable for a particular system, equilibrium relationships are obtained from vapor pressure data using Raoult's law and Dalton's law, where ideal behavior may be assumed. Recently, simulation software including Chemcad, Aspen plus, HYSYS, and UniSim (Honeywell) have incorporated thermodynamic properties in their workbench and thus

allowed phase diagrams of mixtures (T-x-y, x-y, etc) to be readily generated. Various VLE diagrams of mixtures have also been constructed using equilibrium data from Perry's Chemical Engineers' handbook [245] and chemical engineering thermodynamic texts [247, 248, 249 and 294]. These phase diagrams can be accessed in this volume, and are available from the Elsevier companion web site. Figure 10-7 shows the VLE (x-y) diagram for methanol and water at one atmosphere, generated from the T-x-y data of Figure 10-6a. Figure 10-8 shows the enthalpy composition diagram for ethanol-water; Figure 10-8a shows the temperature composition diagram of ethanol-water with a minimum boiling azeotrope at 78.2°C, and Figure 10-7 forms the basis of the McCabeThiele diagram for calculation of continuous fractionation of binary mixtures; this will be presented in the graphical design of a distillation column.

Figure 10-7

VLE diagram for the mixture of methanol and water at 1 atm.

Figure 10-8

Enthalpy-composition diagram for ethanol-water at a pressure of 1 kg/cm^2

(Source: Bosnjakovic, Technishche Thermodynamik, T. Steinkopff, Leipzig, 1935).

Figure 10-8a

Temperature composition diagram for ehtanol and water mixture at 1.013

kPa.00

The distillation process depends on the differences between the compositions of liquid and vapor phases of a mixture at equilibrium. This implies that the temperatures and pressures of the phases must be the same, and no composition change occurs over time. Equilibrium can be achieved after a long period of thorough mixing and contact between the two phases.

An ideal gas is by definition one that follows the ideal gas laws:

$$PV = nRT \tag{10-1}$$

where

P = absolute pressure
V = total volume
n = number of moles of gas
R = ideal gas constant
T = absolute temperature

The conditions under which a given component or mixture approaches ideal behavior depend on its critical temperature and critical pressure. Other applicable principles of gas behavior are Dalton's law of additive pressures and Amagat's law of additive volumes. These are:

$$\pi = p_A + p_B + \ldots \tag{10-2}$$

and

$$V_T = V_A + V_B + \ldots.. \tag{10-3}$$

where

π = total system pressure
V_A = volume of component A
V_B = volume of component B
V_T = total volume of system
p_A = partial pressure of component A
p_B = partial pressure of component B
A, B = components

These laws are only applicable when conditions are such that each component and the mixture obey the ideal gas law. The equilibrium between two phases can be related to the equality of the chemical potential, and defined in terms of the Gibbs free energy as:

$$\mu_i = \left[\frac{\partial G}{\partial \mu_i}\right]_{T,P,n_i} = \overline{G}_i \tag{10-4}$$

where $\overline{G}_i$ is the partial molar Gibbs free energy. From the fundamental thermodynamic relation:

$$dG = V\,dP - S\,dT \tag{10-5}$$

The effect of pressure is mostly observed at constant temperature, as defined by:

$$dG = VdP(\text{constant } T) \tag{10-6}$$

where G is the molar Gibbs free energy and V is the molar volume. For a pure fluid of component i, substitution of Equation 10-1 into Equation 10-6 yields,

$$dG = RT\frac{dP}{P} = RT\ d(\ln P) \tag{10-7}$$

The free energy from Equation 10-7 is restricted to ideal gases, but can be modified to cover real fluids by introducing a new function, the fugacity. Equation 10-7 becomes:

$$dG_i = RT\ d(\ln f_i) \tag{10-8}$$

where f_i is the fugacity of pure i with units of pressure. The fugacity becomes equal to the pressure as the pressure approaches zero.

Thus:

$$\lim_{P \to 0} \frac{f_i}{P} = 1.0 \tag{10-9}$$

For a single component in a mixture at constant temperature, the fugacity is defined by:

$$d\overline{G}_i = RT\ d(\ln \bar{f}_i) \tag{10-10}$$

and:

$$\lim_{P \to 0} \frac{\bar{f}_i}{y_i P} = 1.0 \tag{10-11}$$

If a liquid mixture at temperature T and pressure P is in equilibrium with a vapor mixture at the same temperature and pressure, therefore at an equilibrium condition, with the Gibbs free energy minimized, the thermodynamic criterion will be:

$$\hat{f}_i^V = \hat{f}_i^L \quad (i = 1, 2 \ldots\ldots\ldots N) \tag{10-12}$$

The vapor fugacity can be regarded as a corrected partial pressure, and given by:

$$\hat{f}_i^V = \phi_i^V (P\, y_i) \tag{10-13}$$

where

$\hat{f}_i^V$ = fugacity of component i in the vapor phase

ϕ_i^V = the vapor fugacity coefficient of species i, based on P, T, y_1, y_2.

N = number of chemical species

y_i = mole fraction of species i in the vapor phase.

P = system total pressure.

The fugacity of a component in a mixture depends on its temperature, pressure and composition. The fugacity $\hat{f}_i^L$, of any component i, in the liquid phase is related to the composition (mole fraction) of the component in that phase by:

$$\hat{f}_i^L = \gamma_i\, x_i\, f_i^{oL} \tag{10-14}$$

Similarly, the liquid fugacity can be regarded as a corrected vapor pressure by:

$$\hat{f}_i^L = \phi_i^L\, \gamma_i\, \Psi_i (x_i\, p_i^o) \tag{10-15}$$

where

ϕ_i^L = the liquid fugacity coefficient of species i based on T and p_i^o for pure liquids

Ψ_i = the Poynting correction factor for component i accounting for the effect of pressure on liquid phase fugacity

γ_i = the liquid phase activity coefficient of species i

x_i = mole fraction of species i in the liquid phase

$\hat{f}_i^L$ = fugacity of species i in the liquid phase

f_i^{oL} = the fugacity of species i in the reference (or standard state)

The fugacity of pure liquid i at temperature T and pressure P is given by:

(10-16)

$$f_i^{oL} = P_{vp,i}(T)\,\phi_i^s(T)\exp\left[\int_{P_{vp,i}}^{P} \frac{V_i^L(T,P)}{RT}\,dP.\right]$$

where

$P_{vp,i}$ = the vapor pressure of component i

ϕ_i^s = the vapor phase fugacity coefficient of pure saturated vapor of component i

The exponential term in Equation 10-16 is a correction factor for the effect of pressure on liquid-phase fugacity and is termed the **Poynting correction factor**. In Equation 10-16, V_i^L can be replaced by the partial molar volume of component i in the liquid solution for greater accuracy. For low to moderate pressure, V_i^L is assumed as the saturated liquid molar volume at the specified temperature. Equation 10-16 is simplified to give:

(10-17)

$$f_i^{oL} = P_{vp,i}\,\phi_i^s \exp\left[\frac{V_i^{Ls}\left(P - P_{vp,i}\right)}{RT}\right]$$

Equation 10-17 is used to calculate the reference state fugacity of liquids. Any equation of state can be used to determine ϕ_i^s. For low to moderate pressures, the virial equation is the simplest to use. The fugacities of pure gases and gas mixtures are required for estimation of many thermodynamic properties, *e.g.* entropy, enthalpy and fluid phase equilibria. For pure gases, the fugacity is:

(10-18)

$$\ln f = \frac{1}{RT}\int_0^P V\,dP$$

The fugacity can be calculated from Equation 10-18 once the P-V-T behavior of the fluid is known from an equation of state. The fugacity coefficient is the ratio of the fugacity of a substance to its pressure. For a pure substance:

(10-19)

$$\phi = \frac{f}{P}$$

For the vapor phase, the composition is generally expressed by the mole fraction, y_i:

(10-20)

$$\phi_i = \frac{f_i^V}{y_i\,P}$$

For a mixture of ideal gases, $\phi_i = 1$. The fugacity coefficient, ϕ_i, depends on temperature and pressure. The fugacity coefficient is normalized, such that as $P \to 0$, $\phi_i \to 1$ for all components. At low

pressure, ϕ_i is assumed to be unity. When the relationship between volume and pressure is known, either from direct measurement or from an empirical equation of state, changes in fugacity may be found by integration. Thus:

$$\ln \frac{f_2}{f_1} = \frac{1}{RT} \int_{P_1}^{P_2} V dP \tag{10-21}$$

Since $\phi = 1$ when $P = 0$, absolute values of the fugacity coefficient can be determined. Therefore:

$$\ln \phi = \ln \frac{f}{P} = \frac{1}{RT} \int_0^P \left(V - \frac{RT}{P} \right) dP = \int_0^P (Z - 1) \frac{dP}{P} \tag{10-22}$$

Methods for estimating fugacities of liquid and vapor phases are given by Reid *et al.* [246] and by Walas [247] and other thermodynamic texts [248, 249]. These also give detailed derivations of the above equations and their relationships. The various coefficients in these equations are illustrated below [94]:

ϕ_i^V	The vapor fugacity coefficient accounts for the effect of vapor nonideality on vapor fugacity. It is usually estimated from an equation of state and is based on system temperature, pressure, and vapor mole fraction.
ϕ_i^L	The liquid fugacity coefficient accounts for the effect of vapor nonideality on liquid fugacity. It is estimated in a similar manner to the vapor fugacity coefficient, but it is based on the system temperature and the pure component vapor pressure.
Ψ_i	The Poynting correction factor accounts for the effect of pressure on liquid fugacity. Since ϕ_i^L is evaluated at the vapor pressure of the pure component, Ψ_i is used to account for the difference between the pure component vapor pressure and the mixture pressure. This effect is small and can be neglected at low pressures, but is important at high pressures.
	The liquid activity coefficient corrects the liquid fugacity for the effect of

γ_i	composition. Its value depends on how similar the components are. For two similar components, such as an isobutane-normal butane mixture, the liquid activity coefficient is close to unity. If the components are different, activity coefficients deviate from unity.

10.3. Activity Coefficients

Dissimilar molecules repel each other resulting in activity coefficients (γ_i) greater than 1. So, in a mixture of unlike molecules (e.g. polar water molecules are mixed with nonpolar organic molecules) greater partial pressure is exerted, resulting in a positive deviation from ideality. Figures 10-9a and 10-9b show typical plots of activity coefficients γ_1 and γ_2 of the two components of such a mixture.

Figure 10-9a–b Composition dependence of activity coefficient showing positive deviation from ideality.

At $x_1 = 0$; the pure heavy component (component 2) has an activity coefficient (γ_2) of unity. The activity coefficient of the light component (component 1) will be large and is termed the infinite dilution activity coefficient γ_1^∞.

At $x_1 = 1$, the activity coefficient (γ_1) of light component (component

1) is unity. The activity coefficient of heavy component (component 2) is the infinite dilution activity coefficient γ_2^{∞}.

In the opposite case, where the molecules of the components attract each other, the activity coefficients will be less than 1 as the molecules will exert a lower partial pressure in a mixture than if they were pure. For example, when nitric acid is mixed with water, a lower pressure is exerted. Figures 10-9c and 10-9d show a typical activity coefficient dependence on composition for this case of negative deviation from ideality.

Figure 10-9c–d Composition dependence of activity coefficient showing negative deviation from ideality.

10.4. Excess Gibbs Energy-GE

The excess Gibbs energy $\overline{G}_i^E$ can be expressed by

(10-23)

$$\overline{G}_i^E = RT \ln \frac{\hat{f}_i}{x_i f_i}$$

For an ideal solution, the fugacity of each species is proportional to its mole fraction, i.e.:

(10-24)

$$\hat{f}_i = x_i f_i'$$

where the proportionality constant is the fugacity of pure species i in the same physical state as the solution and at the same temperature, T and pressure, P. Equation (10-24) is known as the ***Lewis/Randall rule*** and it applies to each species in an ideal solution at all conditions of temperature, pressure and composition. The dimensionless ratio $\hat{f}_i / x_i f_i$ is the activity coefficient of species i in solution and is represented by γ_i. This is expressed by:

$$\gamma_i \equiv \frac{\hat{f}_i}{x_i f_i} \tag{10-25}$$

Substituting Equation (10-25) into Equation (10-23) gives:

$$\overline{G}_i^E = RT \ln \gamma_i \tag{10-26}$$

These equations establish a thermodynamic relationship between the excess Gibbs energy and activity coefficient. For a binary system, the excess Gibbs energy is expressed by:

$$\frac{G^E}{RT} = x_1 \ln \gamma_1 + x_2 \ln \gamma_2 \tag{10-27}$$

Table 10-1 shows the vapor-liquid equilibrium data for a methyl ethyl ketone (1) and toluene (2) mixture at 50°C, and Figures 10-10a, 10-10b, 10-10c and 10-10d respectively show plots of pressure against compositions of methyl ethyl ketone (1) in liquid and vapor phases, fugacity against the mole fraction of methyl ethyl ketone (1) in liquid phase and the liquid phase properties of methyl ethyl ketone (1) and toluene (2) mixture. These plots use the experimental data of Pena *et*

al. [250] and have been developed using Excel spreadsheet (Phase-Equil.xls). The detailed explanation and derivation of GE and activity coefficient γ_i are given in standard chemical engineering thermodynamic texts [247, 248].

Table 10-1 VLE data for Methyl Ethyl Ketone (1)/Toluene (2) at 50°C.
Source: M. Díaz Peña, A. Crespo Colin and A. Compostizo, J. Chem. Thermodyn, vol. 10, pp 337–341, 1978.

P/kPa	x_1	y_1	$\hat{f}_1 = y_1 P$	$\hat{f}_2 = y_2 P$	γ_1	γ_2
12.30 (P_2^{sat})	0.0000	0.0000	0.000	12.300		1.000
15.51	0.0895	0.2716	4.212	11.298	1.304	1.099
18.61	0.1981	0.4565	8.496	10.114	1.188	1.026
21.63	0.3193	0.5934	12.835	8.795	1.114	1.050
24.01	0.4232	0.6815	16.363	7.697	1.071	1.078
25.92	0.5119	0.7440	19.284	6.636	1.044	1.105
27.96	0.6096	0.8050	22.508	5.542	1.023	1.135
30.12	0.7135	0.8639	26.021	4.099	1.010	1.163
31.75	0.7934	0.9048	28.727	3.023	1.003	1.189
34.15	0.9102	0.9590	32.750	1.400	0.997	1.268
36.09 (P_1^{sat})	1.0000	1.0000	36.090	0.000	1.000	

Figure 10-10

a. Phase diagram for methyl ethyl ketone (1) and toluence (2) mixture at 50°C. b. Fugacities of methyl ethyl ketone (1) and toluence (2) mixture at 50°C. c. Logarithms of the activity coefficients for methyl ethyl ketone (1) and toluene (2) at 50°C. d. Liquid phase properties for methyl ethyl ketone (1) and toluene (2) mixture at 50°C.

10.5. K-value

The equilibrium constant K can be defined as:

$$K_i = \frac{\text{Mole fraction of component i in vapor phase}}{\text{Mole fraction of component i in liquid phase}} \quad (10\text{-}28)$$

where the K-value is a measure of the tendency of component i to vaporize. It is a function of temperature, pressure and composition. At equilibrium, whenever two of these three variables are fixed, the third variable is also fixed. Therefore, the K value can be defined as a function of composition and pressure; composition and temperature or

temperature and pressure. If the K-value is above 1.0, the component tends to concentrate in the vapor phase; correspondingly, below 1.0, the component tends to concentrate in the liquid phase. If the K-value is unity, the component concentration is equal in the vapor and liquid phases.

Combining Equations (10-12)(10-14)(10-15) and (10-28)

(10-29)

$$K_i = \frac{y_i}{x_i} = \frac{p_i^o}{P}\left[\frac{\phi_i^L \gamma_i \Psi_i}{\phi_i^V}\right]$$

At reasonably low pressures,

(10-30)

$$\frac{\phi_i^L}{\phi_i^V} \approx 1, \text{and } [\gamma_i \ \Psi_i] \approx 1$$

Equation 10-29 becomes:

(10-31)

$$K_i = \frac{p_i^o}{P} \text{ or } y_i P = x_i p_i^o$$

10.6. Ideal Systems

The separation performance of these systems (generally under conditions of low pressure, not close to critical point, and with similar components) can be predicted by Raoult's Law, applying to vapor and liquid in equilibrium. When one liquid is totally miscible in another, the partial pressure of each is decreased. Raoult's Law states that for any mixture, the partial pressure of any component will equal the vapor pressure of that component in the pure state multiplied by its mole fraction in the liquid mixture.

(10-32)

(10-32)

$$p_i^* = p_i x_i = P_i x_i$$

(10-33)

$$p_{ii} = P_{ii}\, x_{ii} \text{(for a second component, ii, in the system)}$$

where

p_i = partial pressure, absolute, of one component in the liquid mixture

x_i = mol fraction of component, i, in the liquid mixture

$P_i^{\times} = P_i$ = vapor pressure of component, i, in its pure state; $P^{\times}_{ii}$ similar by analogy

Many mixtures of liquids do not follow Raoult's Law, since this represents the performance of ideal mixtures. For systems which do follow the ideal gas law and Raoult's Law for the liquid, for each component,

(10-34)

$$y_i = \frac{P_i}{\pi} = \frac{p_i^* x_i}{\pi}$$

(Raoult's Law combined with Dalton's Law)

y_i = mol fraction of component, i, in vapor

π = system total pressure

Raoult's Law is not applicable as conditions approach critical, and for hydrocarbon mixtures accuracy is lost above about 60 psig [81]. Dalton's Law relates composition of the vapor phase to the pressure and temperature well below the critical pressure, i.e., total pressure of a system is the sum of the partial pressures of its components:

(10-2)

$$\pi = p_1 + p_2 + p_3 + \ldots$$

where

p_1, p_2, …= partial pressures of components, 1, 2, …

Therefore, for Raoult's and Dalton's Laws to apply, the relationship between the vapor and liquid compositions for a given component of a mixture is only a function of pressure and temperature, and is independent of the other components in the mixture.

For a binary system, Raoult's law states:

$$p_A = x_A \, p_A^o \tag{10-35}$$

where x_A is the mole fraction of component A, and p_A^o is the vapor pressure of component A.

For the vapor phase, Dalton's law can be expressed as:

$$p_A = \pi \, y_A \tag{10-36}$$

where y_A is the mole fraction of component A. Combining Equation 10-35 and Equation 10-36 gives:

$$x_A \, p_A^o = \pi \, y_A \tag{10-37}$$

$$K = \frac{y_A}{x_A} = \frac{p_A^o}{\pi} \tag{10-38}$$

where K is the equilibrium constant.

Few multicomponent systems exist for which completely generalized equilibrium data are available. The most widely available data are

those for vapor-liquid systems, frequently referred to as vapor-liquid equilibrium distribution coefficients, or the K value. The K values vary with temperature and pressure, and a selectivity that is equal to the ratio of the K values is used. For vapor-liquid systems, this is referred to as the relative volatility, α. For a binary system:

(10-39)

$$\alpha_{A-B} = \frac{K_A}{K_B}$$

where

(10-40)

$$K_A = y_A/x_A$$

and

(10-41)

$$K_B = y_B/x_B$$

(10-42)

$$\alpha_{A-B} = \frac{y_A \cdot x_B}{x_A \cdot y_B}$$

For binary components:

(10-43)

$$x_A + x_B = 1$$

or

(10-44)

$$x_B = 1 - x_A$$

and

(10-45)

$$y_A + y_B = 1$$

or

$$y_B = 1 - y_A \tag{10-46}$$

Substituting Equations (10-44) and (10-46) into Equation (10-42) gives:

$$\alpha_{A-B} = \frac{y_A\,(1 - x_A)}{x_A(1 - y_A)} \tag{10-47}$$

Rearranging Equation (10-47) in terms of y_A gives:

$$y_A = \frac{\alpha_{A-B}\,x_A}{[1 + (\alpha_{A-B} - 1)\,x_A]} \tag{10-48}$$

Equation 10-48 relates equilibrium compositions x_A and y_A in terms of the relative volatility. If it is assumed that α is independent of temperature and composition, Equation 10-48 becomes the equation of the equilibrium line or the x-y curve shown in Figure 10-7. This can be used to determine the compositions of vapor and liquid in equilibrium.

10.7. Henry's Law

Henry's Law applies to the vapor pressure of the solute in dilute solutions, and is a modification or Raoult's Law.

This can be stated as:

$$p_i = kx_i \tag{10-49}$$

where

p_i = partial pressure of the solute

x_i = mol fraction solute in solution

k = experimentally determined Henry's constant

Referring to Figure 10-5, Henry's Law would usually be expected to apply on the vaporization curve for about the first 1 in. of length, starting with zero, *i.e.* over the dilute end, whereas whereas Raoult's Law applies to the upper end of the curve.

Carroll [82] discusses Henry's Law in detail and explains its limitations. This constant is a function of the solute-solvent pair and temperature, but not the pressure, since strictly it is only valid at infinite dilution.

10.7.1. Strict Henry's Law

It is equal to the reference fugacity only at infinite dilution. From [82]:

$$x_i H_{ij} = y_i P \tag{10-50}$$

for restrictions of: $x_i < 0.01$ and $P < 200$ kPa

10.7.2. Simple Henry's Law

This can be stated as:

$$x_i H_{ij} = P \tag{10-51}$$

for restrictions of: $x_i < 0.01$, $y_i = 0$, and $P < 200$ kPa

(10-52)

$$K = \frac{y_i}{x_i} = \frac{p_i^*}{p}$$

where

H_{ij} = Henry's constant

x_i = mol fraction of solute component, i, in liquid

P = pressure, absolute

y_i = mol fraction of solute component, i, in vapor

y_j = mol fraction solvent component, j, in vapor

kPa = metric pressure

Care must be exercised that the appropriate assumptions are made in the evaluation. This requires experience and/or experimentation. Carroll [83] presents evaluation of Henry's Law constant for several multicomponent mixtures; (1) a nonvolatile substance (such as a solid) dissolved in a solvent, (2) a gas in a solution of aqueous electrolytes, (3) mixed electrolytes, (4) mixed solvents, i.e., a gas in equilibrium with a solvent composed of two or more components, (5) two or more gaseous solutes in equilibrium with a single solvent, (6) complex, simultaneous phase and chemical equilibrium.

10.8. K-Factor Hydrocarbon Equilibrium Charts

Hydrocarbon systems are amongst those for which most data have been developed. See following paragraph on K-factor charts. For systems of chemical components where such factors are not developed, the basic relation is:

$$K_i = \frac{y_i}{x_i} = \frac{\gamma_i}{\phi_i}\frac{p_i^*}{P} \qquad (10\text{-}53)$$

For ideal systems: $\gamma_i = \phi_i$

where

K_i = mol fraction of component, i, in vapor phase in equilibrium divided by mol fraction of component, i, in liquid phase in equilibrium

K_i = equilibrium distribution coefficient for system's component, i

p_i^* = vapor pressure of component, i, at temperature

P = total pressure of system = π

γ = activity coefficient

ϕ = fugacity coefficient

The concept of ideality is usually a good approximation for close boiling components of a system, where the components are all of the same “family” of hydrocarbons or chemicals; for example a mixture of paraffin hydrocarbons. When “odd” or non-family components are present, or if the system has a wide boiling range the probability of deviations from nonideality becomes greater.

Ideal conditions are often assumed for preliminary calculations, followed by more rigorous design methods. These may be completely satisfactory, particularly when the activities of the individual components approximate to 1.0. Many process components do not however conform to the ideal gas laws, and erroneous results will be obtained if this behavior is assumed to hold. Errors are not serious when the deviation from ideal is not significant. When data are available to confirm the ideality or nonideality of a system, then the choice of approach is much more straightforward.

K-factors for vapor–liquid equilibrium ratios are usually associated with various hydrocarbons in the presence of some common impurities as nitrogen, carbon dioxide, and hydrogen sulfide [48]. The K-factor is (Equation 10-40) the equilibrium ratio of the mole fraction

of a component in the vapor phase divided by the mole fraction of the same component in the liquid phase. K is generally considered a function of the mixture composition, in which a specific component occurs, plus the temperature and pressure of the system at equilibrium.

The Gas Processors Suppliers Association [79] provides a detailed background development of the K-factors and the use of convergence pressure (see Figure 10-11aFigure 10-11bFigure 10-11c and Figure 10-11d), *i.e.* the pressure of system at a temperature when vapor-liquid separation is no longer possible. Convergence pressure alone does not represent a system's composition effects in hydrocarbon mixtures, but the concept does provide a rather rapid approach for systems calculations and is used for many industrial calculations. These are not well adapted for very low temperature separation systems. Figure 10-11d shows the chart of equilibrium constant for a methane – ethane binary system.

Figure 10-11a Convergence pressures for hydrocarbons (critical locus).

Used permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

Figure 10-11b Pressure vs. K for methane (CH_4) at convergence pressure of 3000 psia (20,700 kPa).

Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

Figure 10-11c Pressure vs. K for n-pentane (n-C_5H_{12}) at convergence pressure of 3000 psia (20,700 kPa).

Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

Figure 10-11d Pressure vs. K for methane-ethane binary.

Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

Charts of reference [79] are for binary systems unless noted otherwise. The convergence pressure method can usually predict equilibrium composition with reasonable accuracy for vapor and liquid phases around the convergence pressure and corresponding temperature. The convergence pressure is that pressure where all proportion of vapor and liquid focus to a point (converges). Generally, vapor and liquid cannot coexist at a pressure higher than the convergence pressure. The K values of the components (binary) approach 1.0. Typically, the convergence pressure is higher than the critical pressure of either of the pure components [295].

The convergence pressure is generally a function of the liquid phase, and assumes that the liquid composition is known from a flash calculation using a first estimate for convergence pressure, and is usually the critical pressure of a system at a given temperature. The

following procedure is recommended by Reference 79:

Step 1. Assume the liquid phase composition or make an approximation. (If there is no guide, use the total feed composition).

Step 2. Identify the lightest hydrocarbon component that is present in at least 0.1 mole % in the liquid phase.

Step 3. Calculate the weight average critical temperature and critical pressure for the remaining heavier components to form a pseudo binary system. (A shortcut approach suitable for most hydrocarbon systems is to calculate the weight average T_c only).

Step 4. Trace the critical locus of the binary mixture consisting of the light component and pseudo heavy component. When the averaged pseudo heavy component is between two real hydrocarbons, an interpolation of the two critical loci must be made.

Step 5. Read the convergence pressure (ordinate) at the temperature (abscissa) corresponding to that of the desired flash conditions, from Figure 10-11a [79].

Step 6. Using the convergence pressure determined in Step 5, together with the system temperature and system pressure, obtain K-values for the components from the appropriate convergence-pressure K-charts.

Step 7. Make a flash calculation with the feed composition and the K-values from Step 6.

Step 8. Repeat Steps 2 through 7 until the assumed and calculated convergence pressures agree within an acceptable tolerance.

For n-pentane at convergence pressure of 3,000 psia (nearest chart), the K-value reads 0.99 (see Figure 10-11c) and other K-values are shown in Appendix J. The DePriester charts [80] provide a reasonable check on this (Figure 10-12aFigure 10-12bFigure 10-12c and Figure

10-12d). Interpolation between charts of convergence pressure can be calculated, depending on how close the operating pressure is to the convergence pressure. The K-factor charts (or K-values) do not change rapidly with convergence pressure, P_k (psia) [79].

Figure 10-12a DePriester Charts; K-Values of light hydrocarbon systems, generalized correlations, low-temperature range.

Used by permission, DePriester, C.L., The American Institute of Chemical Engineers, Chemical Eng. Prog. Ser., 49 No. 7 (1953), all rights reserved.

Figure 10-12b DePriester Charts; K-Values of light hydrocarbon systems, generalized correlations, high-temperature range.

Used by permission, DePriester, C.L., The American Institute of Chemical Engineers, Chemical Eng. Prog. Ser., 49 No. 7 (1953), all rights reserved.

Figure 10-12c Modified DePriester Chart (in S.I. units) at low temperature

(D.B. Dadyburjor, Chem. Eng. Prog., 85, April 1978; copyright 1978, AIChE; reproduced by permission of the American Institute of Chemical Engineers).

Figure 10-12d Modified DePriester Chart (in S.I. units) at high temperature

(D.B. Dadyburjor, Chem. Eng. Prog., 85, April 1978; copyright 1978, AIChE; reproduced by permission of the American Institute of Chemical Engineers).

For light hydrocarbons, approximate K values can be determined from the nomographs prepared by DePriester, and as illustrated in Figure 10-12aFigure 10-12bFigure 10-12c and Figure 10-12d for different temperature ranges. If temperature and/or pressure of the equilibrium mixture are unknown, a trial-and-error method is required. Perry and Green [245], and Smith and Van Ness [248] provide DePriester charts in alternative temperature and pressure units. McWilliams [251] has developed an equation to fit DePriester charts represented by:

(10-54)

where

a_{T1}, a_{T2}, a_{T6}, a_{p1}, a_{p2}, a_{p3} = constants
K = equilibrium constant
p = absolute pressure, psia
T = absolute temperature (°R = °F + 460)

Equation 10-54 is valid from 365.7°R (−70°C) to 851.7°R (200°C) and from pressures from 14.69 psia (101.3 kPa) to 870.1 psia (6000 kPa). If K and p are known, Equation 10-54 can be solved for T. The relationship values can be easily incorporated into a computer program and can be determine more accurate than from the charts. Table 10-2 shows the constants for various components, and Equation 10-54 can be simplified for all components, except n-octane and n-decane. This is expressed by:

(10-55)

Table 10-2 Constants for fit to K Values Using Equation 10-54

Source: Mew Williams, M. L., "An Equation to Relate K-Factors to Pressure a Temperature", Chem. Eng., 80 (25), 138, Oct 29, 1973.

Compound	a_{T1}	a_{T2}	a_{T6}	a_{p1}	a_{p2}	a_{p3}
Methane	−292,860	0	8.2445	−0.8951	59.8465	0
Ethylene	−600,076.875	0	7.90595	−0.84677	42.94594	0
Ethane	−687,248.25	0	7.90694	−0.88600	49.02654	0
Propylene	−923,484.6875	0	7.71725	−0.87871	47.67624	
Propane	−970,688.5625	0	7.15059	−0.76984	0	6.9022
Isobutane	−1,166,846	0	7.72668	−0.92213	0	0
n-Butane	−1,280,557	0	7.94986	−0.96455	0	0
Isopentane	−1,481,583	0	7.58071	−0.93159	0	0
n-Pentane	−1,524,891	0	7.33129	−0.89143	0	0
n-Hexane	−1,778,901	0	6.96783	−0.84634	0	0
n-Heptane	−2,013,803	0	6.52914	−0.79543	0	0
n-Octane	0	−7646.81641	12.48547	−0.73152	0	0
n-Nonane	−2,551,040	0	5.69313	−0.67818	0	0
n-Decane	0	−9760.45703	13.80354	−0.71470	0	0
Note: T is in °R, and p is in psia.						

Solving for T from Equation 10-55 gives:

(10-56)

K values are used along with stoichiometric equations which state that the sum of the mole fractions in liquid and vapor phases must equal to

unity as defined by Equations (10-43) and (10-45) respectively. For N number of components, Equations (10-43) and (10-45) can be expressed by:

(10-57)

where

N = number of components

If only one component is present, then x = 1.0 and y = 1.0, which implies that . This is a simple way of determining the boiling point of a pure compound at any pressure. For example, to determine the boiling point of propane at p = 200 psia, place a straightedge on p = 200 and K = 1.0 on the propane scale on Figure 10-12a. The temperature on the right side of the scale is T = 100°F (560°R), which is the boiling point.

Using Equation 10-54 and values from Table 10-2 gives:

(10-54)

and

This gives a 1.03% deviation from DePriester chart.

The K-factor charts represent pure components and pseudo binary systems of a light hydrocarbon plus a calculated pseudo heavy component in a mixture, when several components are present. For these more complex mixtures it is necessary to determine the average molecular weight of the system on a methane-free basis, and then interpolate the K-value between the two binaries whose heavy component lies on either side of the pseudo-components. If nitrogen is at more than 3–5 mol%, the accuracy becomes poor. Reference 79 provides a more detailed explanation and a more complete set of charts.

10.9. NonIdeal Systems

A system is nonideal because the interaction between molecules alters its behavior. The following groups all contribute to nonideality; hydroxyl groups (-OH), ketone groups (-C=O), aldehyde groups (CHO), halogens (-Cl, -Br) and carboxylic acid groups (-COOH). Systems of two or more hydrocarbon, chemical and water components may be nonideal for a variety of reasons. Accurate experimental data are necessary to predict the distillation performance of these systems. The use of empirical relationships to predict vapor pressure/concentration relationships at specific temperatures and pressures is inaccurate at best.

Prausnitz [54] presents a thorough analysis of the application of empirical techniques in the absence of experimental data. The basic question is to determine accurately the distribution of the respective system components between liquid and gaseous phases. The concepts of fugacity and activity are fundamental to this. For a pure ideal gas the fugacity is equal to the pressure, and for a component, i, in a mixture of ideal gases it is equal to its partial pressure y_i P, where P is

the system pressure. As the system pressure approaches zero, the fugacity approaches ideal. For many systems, the deviations from unity are minor at system pressures less than 25 psig.

The ratio f/fo is termed activity, a. [**Note**: This is not the activity coefficient.] The activity is an indication of how “active” a substance is relative to its standard state (not necessarily zero pressure), fo. The standard state is the reference condition; most references are to constant temperature, with composition and pressure varying as required. Fugacity becomes a corrected pressure, representing a specific component's deviation from ideal. The fugacity coefficient is:

(10-58)

The Virial Equation of State for gases is generally:

(10-59)

where

B, C, D, *etc.* = Virial coefficients, independent of pressure or density, and for pure components are functions of temperature only

V = molar volume

Z = compressibility factor

Fugacities and activities can be determined using this concept. Other important equations of state, which can be related to fugacity and activity have been developed by Redlich-Kwong [56] with Chueh [10], which develops the original Redlich-Kwong equation, and Palmer's summary of activity coefficient methods [51].

Activity coefficients are equal to 1.0 for an ideal solution when the mole fraction is equal to the activity. The activity (a) of a component, i, at a specific temperature, pressure and composition is defined as the ratio of the fugacity of i at these conditions to the fugacity of i at the standard state [54].

(10-60)

(Zero superscript indicates a specific pressure and composition)

The activity coefficient γ_i is:

(10-61)

The ideal solution law, Henry's Law, also enters into the establishment of performance of ideal and nonideal solutions.

The Redlich-Kister [55, 57] equations provide a good technique for representing liquid phase activity and classifying solutions.

The Gibbs-Duhem equation allows the determination of activity coefficients for one component from data for those of other components.

Wilson's [77] equation has been found to be quite accurate in predicting the vapor-liquid relationships and activity coefficients for miscible liquid systems. The results can be expanded to as many components in a multicomponent system as necessary without additional data requiring. This makes Renon's and Prausnitz's [58] techniques valuable for multicomponent systems when applying

computational solutions.

Renon's technique is applicable to partially miscible systems as well as those with complete miscibility. This is described in the reference above and in (54).

Other specific techniques are applicable to particular situations; they should be investigated to make sure that they apply well to the system under consideration. They are often expressed in terms of the effective “K” for the components, i, of a system. Frequently used methods are those of Chao-Seader (CS), Peng-Robinson (PR), Renon, Redlich-Kwong, Soave Redlich-Kwong (SRK/RSK), Wilson. Table 10-3 lists the equations of state that are applied most widely.

Table 10-3 Useful Equations of State.

Name	Equation	Equation constants and functions
(1) ideal gas law		None
(2) Generalized		as derived from data
(3) van-der-Waals		a and b are species-dependent constants and estimated from the critical temperature and pressure
(4) Redlich		

Redlich-Kwong (R-K)		
(5) Soave-Redlich-Kwong (S-R-K or R-K-S)		
(6) Peng-Robinson (P-R)		

The van-der-Waals equation (3) in Table 10-3 was the first successful approach to the formulation of an equation of state for a nonideal gas. It is seldom used because its range of application is too narrow. The constants a and b are estimated from the critical temperature T_c and pressure P_c. The generalized equation of state (2) given in Table 10-3 defines the compressibility factor Z, which is a function of reduced pressure P_r, ($P_r = P/P_c$) and reduced temperature T_r, ($T_r = T/T_c$). The critical compressibility factor Z_c, or the acentric factor, ω developed by Pitzer *et al.* [252] is determined from experimental P-V-T data. It accounts for differences in molecular shape, is determined from the vapor pressure curve and defined as:

(10-62)

Poling *et al.* [253] give extensive lists of values of ω. The equation, developed by Redlich and Kwong (4), in Table 10-3 is a considerable improvement over the van der Waals equation. Shah and Thodos [254] showed that a simple R-K equation comparable accuracy with

equations containing many more empirical constants. The R-K equation can also approximate the liquid phase region. A cubic equation in V is obtained if the R-K equation is expanded to obtain a common denominator. Alternatively, (3) and (4) in Table 10-3 can be combined to eliminate V, and hence give the compressibility factor Z form of the R-K equation, as expressed by:

(10-63)

where

(10-64)

(10-65)

Equation 10-63 can be solved analytically for three roots (e.g. see Appendix-G).

10.10. Thermodynamic Simulation Software Programs

One of the basic tests of a thermodynamic model is to find relevant experimental data to test the predictions of the model. The comparison helps to decide whether the model is good or whether another type would be more suitable. Simulation programs have in-built packages which provide accurate thermodynamic property predictions for hydrocarbons and non-hydrocarbon fluids. Many of these simulation programs have databases of over 1500 components and over 10,000 of fitted binary coefficients. If a component cannot be found within the database, the software programs provide a selection of estimation methods for creating appropriate hypothetical compounds (e.g. pseudo

hydrocarbon components of crude oil for determining the true boiling point (TBP) of crude oil). Experimental pure component data can be used as input to the regression package. Alternatively, the designers can supply a set of their own data. The regression package will fit the input data to one of the mathematical expressions, and thus allows the designers to obtain simulation results for specific thermophysical properties which closely match their experimental data.

Selecting the correct thermodynamic models from packages in the simulation programs (e.g. UniSim, Hysys, Aspen, ChemCad, Pro/II, ProSim, etc.) enables the designer to predict properties of mixtures ranging from well-defined light hydrocarbon systems to complex oil mixtures and highly nonideal (non-electrolytic) chemical systems. To obtain a successful simulation, the input parameters and regression model must be carefully chosen. Changes in temperature and pressure can affect the accuracy of a simulation quite drastically if inappropriate choices are made.

The thermodynamic models used to predict the chemical interaction properties are the NRTL, Wilson, SRK, UNIQUAC, and UNIFAC models. These are incorporated in various simulation packages *e.g.* Pro II, HYSYS, Aspen Plus, ProSim Plus, ChemCad and UniSim (Honeywell).

The procedure for selection, as illustrated by Seader and Henley [255], is as follows:

RK = Redlich-Kwong
PR = Peng Robinson
SRK/RSK = Soave Redlich Kwong
NTRL = Non-random two-liquid
UNIQUAC = Universal quasichemical
UNIFAC = UNIQUAC Functional group activity

1. Characterize the mixture by the chemical types present:

Light gases (LG), Hydrocarbons (HC), Polar organic compounds (PC), and aqueous solutions (A), with or without electrolytes (E).

2. If the mixture is (A) with no (PC), and if electrolytes are present, select the modified NRTL equation. Otherwise, select a special model, such as one for sour water (containing NH_3, H_2S, CO_2, etc.) or aqueous amine solutions.

3. If the mixture contains (HC), with or without (LG), and has a wide boiling range, choose the corresponding states method of Lee-Kesler-Plöcker. If the boiling range of a mixture of (HC) is not larger selection depends on pressure and temperature.

4. For all temperatures and pressures, the Peng-Robinson (PR) equation is suitable.

5. For all pressures and non-cryogenic temperatures, the Soave-Redlich-Kwong (SRK/RSK) equation is recommended.

6. For all temperatures, but not pressures in the critical region, the Benedict-Webb-Rubin-Starling method is suitable.

7. If the mixture contains (PC), the selection depends on whether (LG) are present. If they are, the Predictive Soave-Redlich-Kwong (PSRK) model is recommended. If not, then a suitable liquid-phase activity coefficient method should be chosen.

8. If binary interaction coefficients are not available, select the UNIFAC method. Its results should be considered as only a first approximation.

9. If the binary interaction coefficients are available, and splitting into two liquid phases will not occur, select the Wilson or NRTL equation.

10. If phase splitting is likely, select the NRTL or

UNIQUAC model.

Table 10-4 shows the conditions of some of these models, and Figure 10-13 [249] gives the selection process in the form of a flow chart. Table 10-5 lists some typical systems and recommended correlations.

Table 10-4 Approximate Guides for Selection of K-Values Methods.

Source: Wankat, P.C., Separation Process Engineering, 2nd, Prentice Hall, 2007.

Chemical systems		
Low MW alcohol and hydrocarbons	Wilson	
Higher MW alcohol and hydrocarbons	NRTL	
Hydrogen bonding systems	Margules	
Liquid–liquid equilibrium	NRTL/UNIQUAC	
Water as a second liquid phase	NRTL	
Components in a homologous family	UNIQUAC	
Low-pressure systems with associating vapor phase		
Light Hydrocarbon and Oil Systems		
Natural gas systems w/sweet and sour gas	SRK/PR	
Cryogenic systems	SRK/PR	
Refinery mixtures with p < 5000 psia	SRK/PR	
Hydrotreaters and reformers	Grayson-Stread	
Simple paraffinic systems	SRK/PR	
Heavy components w/NBP > 1,000°F	BK10	
Based on polarity and ideality		
Nonpolar – nonpolar	Ideal & nonideal	any activity coefficient model
Nonpolar – weakly polar	Ideal	any activity coefficient model
Nonpolar – weakly polar	Nonideal	UNIQUAC

Nonpolar – strongly polar	Ideal	UNIQUAC
Nonpolar – strongly polar	Nonideal	Wilson
Weakly polar – weakly polar	Ideal	NRTL
Weakly polar – weakly polar	Nonideal	UNIQUAC
Weakly polar – strongly polar	Ideal	NRTL
Weakly polar – strongly polar	Nonideal	UNIQUAC
Strongly polar – strongly polar	Ideal	UNIQUAC
Strongly polar – strongly polar	Nonideal	NO RECOMMENDATION
Aqueous – strongly polar		UNIQUAC
NRTL = non-random two liquid model; SRK = Soave-Redlich-Kwong; PR = Peng-Robinson, BK10 = Braun K10 for petroleum; UNIQUAC = universal quasichemical theory.		

Figure 10-13

Flow chart to select the best thermodynamic model. The abbreviation BIP is used to mean binary interaction parameters

(source: Elliot, J. R., and Carl. T. Lira, Introductory Chemical Engineering Thermodynamics, Prentice Hall Int. Series, 1999).

Table 10-5 Generalized Antoine Constant Functions for the SRK Equation

Source: DeDoes, A. J., et al., CEP, pp 39–44, Jan. 2007.

10.11. Vapor pressure

If the vapor and liquid of a pure component are in *equilibrium*, then the equilibrium pressure is known as the vapor pressure. Vapor pressure can be defined as the pressure that a pure component exerts at a given temperature, when both liquid and vapor phases are present. It is a physical property of a pure chemical component, and depends only on temperature.

Vapor pressure is an essential parameter in the analysis of separation processes, such as flash separation, distillation or absorption. Various approaches such as tables and charts listing vapor pressure or boiling point among their physical properties are used in these analyses. These include steam tables, thermodynamic property tables and phase equilibrium diagrams such as pressure-enthalpy, pressure-volume or detailed some temperature-entropy diagrams.

DeDoes *et al.* [256] have detailed some drawbacks of these approaches. They have also developed a generalized method, which combines pressure-volume-temperature (PVT) equations of state with thermodynamic criteria that are related to chemical potential or fugacity. The method uses three constants; critical temperature, critical pressure and the acentric factor. Other phase properties, such

as molar volume or density, residual enthalpy and residual entropy are also predicted and generate mixture compositions, K-values or relative volatilities when combined with appropriate mixing rules or a liquid solution model. The generalized method uses a three-constant Antoine vapor pressure relationship for any substance for which the critical temperature, critical pressure and acentric factor are known. Vapor pressure predictions have been found to agree to within ±0.1% with those generated by an algorithm-based approach over a range of reduced temperatures between 0.60 to 1.00 for Soave-Redlich-Kwong (SRK) and between 0.7 and 1.00 for Peng-Robinson (PR) as shown in Table 10-6. Figure 10-14 shows Cox chart plots of vapor pressure for various substances, Figure 10-14a and Figure 10-14b show the vapor pressures for light hydrocarbons at a low temperature range and at a high temperature range respectively.

Table 10-6 Generalized Antoine constant functions for the SRK equation.
Source: DeRoes, A. J., et al., CEP, pp 39-44, Jan. 2007.

	= constants in generalized Antoine

equation, f = acentric factor function, T_r = reduced temperature.

Figure 10-14

Cox chart vapor pressure plots.

(Source: A. S. Foust et. al., Principles of Unit Operations, Wiley, New York, p. 550, 1960).

Figure 10-14a Low-Temperature Vapor Pressures for Light Hydrocarbons. Used by permission, Gas Processors Suppliers Association
Data Book, 12th Ed., V. 1 and 2 (2004) Tulsa, Okla.

Figure 10-14b High-Temperature Vapor Pressures for Light Hydrocarbons. Used by permission, Gas Processors Suppliers Association
Data Book, 12th Ed., V. 1 and 2 (2004) Tulsa, Okla.

Vapor pressure determination using the Clausius-Clapeyron and the Antoine Equations

Standard chemical engineering texts [246, 247 and 248] provide equations to predict vapor-pressure of compounds, and a commonly given equation for the vapor pressure (Refer to chapter 3, page 109, volume 1 of this series) is:

(3-16)

(10-65)

where P is the vapor pressure in mm Hg, and A, B, C, D and E are constants. Sometimes it is useful to estimate the saturated vapor pressure from the Clausius-Clapeyron equation as:

(10-66)

and

(10-67)

where is the molar latent heat of vaporization and T is the absolute temperature. is a function of temperature; but if it assumed to be independent of temperature, Equation 10-67 can be integrated to give:

(10-68)

Equation 10-68 has been found to be fairly accurate for correlating the temperature dependence of the vapor pressure of liquids over limited temperature ranges.

Note: should be a linear function of 1/T, where T is the absolute temperature with the lower limit of integration as the normal boiling conditions, *i.e.* at

where is the atmospheric pressure, and R are in a consistent set of units. One can obtain an estimate of the temperature variation of the heat of vaporization by noting that the integration of Equation 10-68 can be carried out as an indefinite rather than definite integral, which gives:

(10-69)

where C is a constant. A plot of versus 1/T, should give a straight line with a slope equal to , if the heat of vaporization is independent of temperature, and a curve if varies with temperature. Equation 10-69 can be re-written as:

(10-70)

with , and it is reasonably accurate for estimating the temperature dependence of the vapor pressure over small temperature ranges.

Correspondingly, the Antoine equation can be expressed by:

(10-71)

or

(10-72)

where A, B and C are dimensional constants, is the vapor pressure and T is the temperature. Equation 10-72 is used to correlate vapor pressure accurately over the range from 1 to 200 kPa. The generalized Antoine equation is expressed as:

(10-73)

where A′, B′ and C′ are constants that depend upon the acentric factor. Table 10-6 and Table 10-7 represent the generalized Antoine constants derived from the vapor pressure predictions of the RSK equation and the PR equation respectively. Table 10-8 shows the transformation required to convert values of A′, B′ and C′ to Antoine constants of A, B and C for use with various temperature and pressure units in Equation 10-70. An Excel spreadsheet program (Example 10-1.xls) has been developed to determine the vapor pressure, molar volumes, and fugacity coefficients from the Soave-Redlich-Kwong (SRK) equation of state. The spreadsheet also compares the vapor pressure obtained by using the generalized Antoine equation with the polynomial expressed in Volume 1 (page 109), and the calculated vapor pressure from SRK equation. Another spreadsheet program (Example 10-2.xls) has been developed to compare the vapor pressure of components using the Antoine Equation 10-71 and coefficients in Table 10-9 with the Clausius-Clapeyron Equation 10-68, over a range of temperatures. The saturated vapor pressure curves of acetone that were predicted by Equations 10-68 and 10-71 are shown in Figure 10-15. The parameters are taken from Reid, *et al.* [246], and have the

values , , . At pressures <10 atm., the Clausius-Clapeyron equation reproduces the experimental data very closely and would be appropriate for use in vapor-liquid equilibrium (VLE) calculations. At high pressure, the predictions are not so good, and by comparison, the Antoine coefficients can only be obtained by regression of experimental vapor pressure data. If the constant ΔH_{vap} in the Clausius-Clapeyron equation is treated as an adjustable parameter and fitted to the experimental vapor pressure data, a value of 7175 cal/gmol produces a curve that fits the data as accurately as the Antoine equation [262].

Table 10-7 Generalized Antoine Constant Functions for the PR Equation.
Source: DeDoes, A. J., et al., CEP, pp 39-44, Jan. 2007.

= constants in generalized Antoine equation, f = acentric factor function, T_r = reduced temperature.	

Table 10-8 Transforming Antoine Constants A′, B′ and C′ from Table 10-6 to Dimensional Constants A, B and C for use in Equation 10-71.
Source: DeDoes, A. J., et al., CEP, pp 39-44, Jan. 2007

	The units of P_c must be absolute pressure Vapor pressure P^* in Equation 10-71 will have the same units as P_c
	The units of T_c must be absolute temperature in K or °R
	Temperature T in Equation 10-71 will have the same as T_c
	Temperature T in Equation 10-71 in °C and T_c in K
	Temperature T in Equation 10-71 in °F and Tc in °R

Table 10-9 Antoine Coefficients for Selected Substances

Source: Vapor-Liquid Equilibrium Data Collection of the DECHEMA Chemistry Data Series, except where noted. The form of the equation is $\log_{10} P^{vap} = A - B/(T + C)$ with pressure in mmHg and temperature in °C.

Substance	A	B	C	Temperature range, °C
Acetone	7.63132	1566.69	273.419	57 to 205
Water	8.01767	1715.70	234.268	100 to 265
Benzene	6.87987	1196.76	219.161	8 to 80
Toluene	6.95087	1342.31	219.187	−27 to 111
Ethylene glycol	8.09083	2088.936	203.454	50 to 150
Hexane	6.91058	1189.64	226.280	−30 to 170
p-Xylene	6.99053	1453.430	215.310	27 to 166
Ethanol	8.11220	1592.864	226.184	20 to 93
Acetic acid	8.02100	1936.010	258.451	18 to 118
Acetaldehyde	7.20812	1099.810	233.945	−82 to 20
Methanol	8.08097	1582.271	239.726	15 to 84
Methyl ethyl ketone	7.06356	1261.340	221.969	43 to 88

Chloroform	6.95465	1170.966	226.232	−10 to 60
Ethylenediamine	8.09831	1893.720	245.676	11 to 117
4-Methyl-2-pentanol	8.46706	2174.869	257.780	25 to 133
Dichloromethane	7.08030	1138.910	231.450	−44 to 59
1,3-Butadiene	6.85364	933.586	239.511	−75 to −2
Styrene	7.50233	1819.810	248.662	−7 to 145
Ethyl acetate	7.10179	1244.950	217.881	16 to 76
Vinylacetylene	7.02515	999.110	235.817	−93 to 5
Acetic anhydride	7.69301	1781.29	230.395	2 to 140
Dichlorosilane[§]	7.18600	1224.50	273.16	−45 to 121
Trichlosilane	6.95524	1102.900	238.865	−81 to 32
Silicon tetrachloride	7.02404	1212.890	235.910	−63 to 57
Hydrogen chloride[†]	7.44899	868.358	274.904	−85 to 36
[§]Bawa, M. S., Texas Instruments (1988).				
[†]Ohe, S., Computer-Aided Data Book of Vapor Pressure, Data Book Publishing Company, Tokyo, Japan (1976).				

Figure 10-15

Vapor pressure of Acetone vs. temperature

Example 10-1: Soave-Redlich-Kwong Prediction of Vapor Pressure

For n-pentane, T_c = 469.76K, P_c = 33.7 bar, ω = 0.252. Use these data to verify that the SRK equation predicts a vapor pressure of 1.05bar at 309.2K. Also, verify these data with the generalized Antoine equation. R = 83.14 cm^3-bar/mol-K.

Solution

The Excel spreadsheet, Example 10-1.xls, has been developed to calculate the molar volumes of liquid and vapor of any component such as pentane, its fugacity coefficient in liquid and vapor phases, and the vapor pressure of n-pentane using the SRK, the generalized Antoine equation and a polynomial Equation (3-61). The percentage deviation in the value of the vapor pressure between the SRK equation and the generalized Antoine's equation can be calculated. Equation 10-73 depends on the reduced temperature and acentric factor, and Table 10-6 provides the constants A′, B′ and C′ for SRK at a range of reduced temperature, T_r . The spreadsheet calculates the vapor pressure of SRK equation by using a Goal seek/Solver optimization tool in the Excel spreadsheet. The SRK (Equation 5) in Table 10-3 is:

(10-74)

where

(10-75)

(10-76)

(10-77)

The constant parameter a_{SRK} in Equation 10-75 is calculated from Equation 10-77, with known and calculated values of the critical

temperature, reduced temperature, critical pressure and the Universal gas constant; the constant parameter b_{SRK} is calculated from Equation 10-76. Equation 10-74 is rearranged in terms of the molar volume as a cubic equation and is solved for liquid and vapor molar volumes using the developed Excel spreadsheet (Example 10-1.xls). The SRK equation is now expressed in terms of the molar volume as:

(10-78)

Using the calculated vapor pressure from the generalized Antoine Equation 10-73, the molar volumes (three real roots) are determined. The liquid molar volume and the vapor molar volume (i.e. the lowest and highest values) are then chosen from these three roots, and are used to determine the fugacity of the liquid and vapor phases from the following equations [256]:

(10-79)

(10-80)

(10-81)

Equation 10-74 is rearranged in the form

(10-82)

Equation 10-82 in the Excel spreadsheet (Example 10-1.xls) is set to zero (cell B44) and is optimized by changing cell G37 using Solver or Go-Seek optimizer in the spreadsheet. This is carried out for known value of the calculated molar volume V, by changing the value of the vapor pressure in the cell. When the final value of Equation 10-82 becomes 0, then the computed value of P (cell G37) gives the required vapor pressure. Since the calculated fugacity coefficients for both liquid and vapor are equal, the vapor pressure as calculated from SRK equation is the estimated known vapor pressure at the specified temperature. In this example, the vapor pressure of n-pentane at 309.2 K is 1.05 bar. Figures 10-16 a-d show snap shots of the Excel spreadsheet for calculating the vapor pressure using the SRK model.

Figure 10-16a–b Snapshot of the Excel spreadsheet for calculating the vapor-pressure using SRK method and Antoine's Equation (Example 10-1).

Example 10-2: Use of the Clausius-Clapeyron Equation to Predict Heat of Vaporization

The vapor pressure of liquid 2, 2, 4 trimethyl pentane at various temperatures is given below. Estimate the heat of vaporization of this compound at 25°C.

Vapor pressure (kPa)	0.667	1.333	2.666	5.333	8.000	13.33	26.66	53.33	101.32
Temperature (°C)	−15.0	−4.3	7.5	20.7	29.1	40.7	58.1	78.0	99.2

Solution

The Excel spreadsheet Example 10-2.xls (vapor-pressure-calculation-2.xls) has been developed to estimate the heat of vaporization and Figure 10-17 shows the trend line from the spreadsheet indicating that is constant over the whole range of temperature. The value of the slope from the trend line is:

Figure 10-17

A plot of ln P^{vap} versus 1/T of 2,2,4 trimethyl pentane.

and the estimated

The calculated heat of vaporization, ΔH_{vap} = 35.16 kJ/mol.

Example 10-3: Construction of Temperature Composition Diagram, Equilibrium Diagram and Plot of Relative Volatility vs Mole Fraction of Phenol in Liquid for Phenol-Orthocresol mixture

The following vapor pressures were obtained for phenol and orthocresol.

Temperature (K)	Vapor pressure (kN/m^2) Orthocresol	Phenol
387.0	7.70	10.0
387.9	7.94	10.4
388.7	8.21	10.8
389.6	8.50	11.2
390.3	8.76	11.6
391.1	9.06	12.0
391.9	9.40	12.4
392.7	9.73	12.9
393.3	10.00	13.3

Assuming Raoult's and Dalton's laws apply; construct the following data for a total pressure of 10 kN/m^2

1. A temperature composition diagram.
2. An x–y diagram.
3. Relative volatility against mole fraction of phenol in liquid.

Solution

If Raoult's law and Dalton's law apply, the following equations may be used.

(10-2)

(10-31)

where

= the total pressure

p_A = the partial pressure of component A
p_B = the partial pressure of component B
= the vapor pressure of component A.
= the vapor pressure of component B.

Hence from Equations (10-2) and (10-31):

(10-83)

(10-84)

The relative volatility, , is defined in terms of the more volatile component (eg. phenol) by:

(10-85)

Table 10-10 shows the results of the computation.

Table 10-10 Mole Fractions x_A, y_A and α_{A-B} in Example 10-2.

Temperature, K	**x_A (from Eq. 10-**	**y_A (from Eq. 10-**	**(from Eq. 10-85)**

	83)	84)	
387.0	1.0	1.0	1.300
387.9	0.837	0.871	1.310
388.7	0.691	0.746	1.315
389.6	0.556	0.622	1.318
390.3	0.437	0.506	1.324
391.1	0.320	0.384	1.325
391.9	0.200	0.248	1.319
392.7	0.085	0.110	1.326
393.3	0	0	1.330

Figure 10-18 and Figure 10-19 show the temperature composition diagram and the equilibrium diagram respectively. The values obtained for the relative volatility vary between 1.300 and 1.330 and the average value of 1.319 could be used in subsequent computations. The Excel spreadsheet Example 10-3.xls shows the calculations of , and .

Figure 10-18

Phase diagram for the mixture phenol and orthocresol at 10 kN/m.^{2}

Figure 10-19

VLE data for the mixture phenol and orthocresol at 10 kN/m.[2]

10.12. Azeotropic Mixtures

Azeotrope is derived from the ancient Greek words meaning "*to boil unchanged*", and is defined as a mixture of composition at which the equilibrium liquid and vapor compositions are equal. Figure 10-5Figure 10-6aFigure 10-6b and Figure 10-7 are phase diagrams for normal systems, showing that the concentration of the less volatile component increases with increase in the dew and bubble points. However, if the components exhibit strong physical and chemical interaction, the resulting phase diagrams are significantly different. In such systems, there is a critical composition (i.e. the point of intersection of the equilibrium curve with the 45° line) for which the vapor and liquid compositions are identical. At this point, the components cannot be separated at the given pressure, and such mixtures are termed azeotropes.

Azeotropes are common when polar organic compounds, such as alcohols, ketones, ethers and acids are mixed with water, and this complicates their separation by distillation. In practice, there will be no change in liquid and vapor compositions from tray to tray, and hence a simple distillation process will not separate the mixture. To determine methods that will separate such mixtures, it is useful to examine their deviations from Raoult's Law.

Azeotropes occur in various systems as binary, ternary and multicomponent mixtures, and can be "homogeneous" (single liquid phase) or "heterogeneous" (two liquid phases). They can exhibit a minimum-boiling point or maximum-boiling point.

A minimum-boiling azeotrope boils at a temperature lower than either of the pure components. Consequently, when distilling a binary system made up of these components, the top is azeotrope and the bottom is one or the other of pure components depending on the "side" of the azeotrope that the feed was. A maximum-boiling azeotrope boils at a temperature higher than either of the pure components and leaves from the bottom of a column and the top product is the high boiling point component containing the More Volatile Component (MVC) at low concentrations. The top product is the MVC when it is present at high concentrations. When liquid phase separation occurs, the boiling temperature of the mixture as well as the vapor phase composition remains constant until one of the liquid phases disappears. Under such conditions, depending on the composition, a mixture of the two liquids will leave the top of the column and either of the components will leave at the bottom.

Azeotropic mixtures consist of two or more components, and are surprisingly common in distillation systems. It is therefore essential with any new system to determine whether an azeotrope exists. Fortunately, if experimental data are unavailable, reference can be made to excellent lists for known azeotropic systems, with vapor pressure information [20, 28, 43]. Typical forms of representation of azeotropic data are shown in Figure 10-20Figure 10-21a and Figure 10-21b respectively. These are homogeneous, with one liquid phase at the azeotrope point. Figure 10-22a and Figure 10-22b illustrate a heterogeneous azeotrope where two liquid phases are in equilibrium with one vapor phase. The system butanol-water is an example of the latter. Chloroform-methanol and acetone-chloroform are examples of binary homogeneous azeotropes with "minimal-boiling point" and "maximum-boiling point" respectively.

Figure 10-20

a. Phase diagram for the mixture of chloroform and methanol at 1 atm. b. VLE diagram for choloroform and methanol mixture at 1 atm.

Figure 10-21a Phase diagram for the mixture chloroform and acetone at 1 atm.

Figure 10-21b Vapore Liquid equilibrium diagram for chloroform acetone mixture at 1 atm.

Figure 10-22a Phase diagram for the mixture of water and 1-butanol at 1 atm.

Figure 10-22b VLE diagram for the mixture of water and 1-butanol at 1 atm.

A “minimum” boiling binary azeotrope exhibits a constant composition as shown by its crossing of the x = y along the 45° line in Figure 10-20 b, which boils at a lower temperature than either of its pure components. This class of azeotrope results from positive deviations from Raoult's Law. Likewise, the “maximum” (Figure 10-21 b) boiling binary azeotrope represents a negative deviation from Raoult's Law, and exhibits a constant boiling point greater than either of the pure components. At the point where the equilibrium curve crosses x = y, 45° line, the composition is constant and cannot be further purified by normal distillation. Both minimum and maximum boiling point azeotropes can be modified by changing the system pressure, and/or addition of a third component, which should be chosen such that will form a minimum boiling azeotrope with one of the original pair. To be effective the new azeotrope should boil well below or above the original azeotrope. Using this technique, one of the original components can often be recovered as a pure product, while still obtaining the second azeotrope for separate purification.

In a minimum-boiling azeotrope, the molecular species repel each other, whereas in a maximum-boiling azeotrope they attract each other. However, if the azeotrope occurs in the composition range in which the two liquids are immiscible, phase splitting occurs and a heterogeneous azeotrope is formed. Table 10-11 lists some examples of binary mixtures that form azeotropes.

Table 10-11 Examples of Binary Mixtures with an Azeotrope

Source: Stichlmair, J. G., and James R. Fair, Distillation: Principles and Practi
Wiley-VCH, 1998

Formula	Name	B.P., °C	Formula	Name	B.P., °C	Azeotropic data		T
						xa, mol %	B.P., °C	

H_2O	water	100	HBr	hydrobromic acid (a)	−67	16.7	126	m
			HCl	hydrochloric acid (a)	−85	11.1	108.6	m
			HNO_3	nitric acid (a)	86	38.3	121	m
			$CHCl_3$	chloroform (a)	61.2	83.9	56.12	h
			CH_2Cl_2	dichloromethane (a)	40.1	93.3	38.1	h
			CH_3NO_2	nitromethane (b)	101.2	51.1	83.6	h
			C_2H_6O	ethanol (a)	78.3	90.4	78.2	m
			$C_3H_6O_2$	methyl acetate (a)	57	87.2	56.4	m
			C_3H_8O	n-propyl alcohol (a)	97.3	42.2	87.8	m
			$C_4H_6O_2$	methyl acrylate (a)	80	72.9	71	m
			C_4H_8O	tetrahydrofurane (a)	66	81.7	64	m
			C_4H_8O	2-butanone (a)	79.6	66.2	73.4	m
			$C_4H_8O_2$	butyraldehyde (a)	74	79.6	68	m
			$C_4H_8O_2$	dioxane (b)	101.3	51.7	87.8	m
			$C_4H_8O_2$	ethyl acetate (a)	77.1	68.2	70.4	h
			$C_4H_{10}O$	butyl alcohol (b)	117.4	75.2	92.7	h
			$C_4H_{10}O_2$	2-ethoxyethanol (b)	135.1	92.5	99.4	m
			C_5H_5N	pyridine (b)	115	76.8	94	m
			$C_5H_{12}O$	ethyl propyl ether (a)	63.6	83.1	59.5	h
			C_6H_6	benzene (a)	80.2	70.4	69.3	h
			C_7H_8	toluene (b)	110.7	55.6	84.1	h
$CHCl_3$	chloroform	61.2	CH_4O	methanol (b)	64.7	65.0	53.4	m

			C_2H_6O	ethanol (b)	78.3	83.7	59.4	m
			C_3H_6O	acetone (a)	56.1	36.7	64.4	m
CH_4O	methanol	64.7	CH_2Cl_2	dichloromethane (a)	40.1	70.5	37.8	m
			C_3H_6O	acetone (a)	56.2	80.2	55.5	m
			C_4H_8O	2-butanone (b)	79.6	84	63.5	m
			$C_4H_8O_2$	ethyl acetate (b)	77.1	69.9	62.3	m
			C_6H_6	benzene (b)	80.2	61	57.5	m
			C_7H_8	toluene (b)	110.7	86.5	63.8	m
			C_7H_{16}	n-heptane (b)	98.5	76.5	59.1	m
			C_8H_{18}	n-octane (b)	125.6	90.2	63	m
C_2H_6O	ethanol	78.3	$C_4H_6O_2$	methylacrylate (b)	80	57.5	73.5	m
			C_4H_8O	2-butanone (b)	79.6	50.0	74	m
			$C_4H_8O_2$	ethylacetate (b)	77	53.8	71.8	m
			$C_4H_{10}O_2$	Eth.methoxymeth (a)	65.9	77	64	m
			$C_5H_{12}O$	ethylpropylether (a)	63.6	61	61.2	m
			C_6H_{14}	n-hexane (a)	68.7	66.5	58.4	m
			C_7H_{18}	toluene (b)	110.7	81	76.7	m
C_3H_6O	acetone	56.2	C_6H_{12}	cyclohexane (b)	80.8	74.6	53	m
			C_6H_{14}	hexane (b)	68.9	68.1	49.8	m
			C_7H_{16}	heptane (b)	98.4	93.6	55.9	m
$C_4H_{10}O_2$	2-ethoxyethanol	135.1	C_8H_{10}	ethylbenzene (b)	136.2	56.9	127.1	m
			C_8H_{18}	octane (a)	125.8	61	116.1	m
C_5H_5N	pyridine	115	C_7H_8	toluene (b)	110.7	75.3	110.1	m
			C_8H_8	n-octane (b)	125.8	<92.9	128.8	m

C_6H_6	benzene	80.2	C_3H_8O	2-propanol (b)	82.5	61	71.8	m
			C_6H_{12}	cyclohexane (b)	80.8	53	77.6	m
B.P. = boiling point, xa = mol % of component a (low boiler) at the azeotrope, p 1.013 bar								

For “minimum” boiling azeotrope, the partial pressures of the components will be greater than predicted by Raoult's Law, and the activity coefficients will be >1.0.

(10-86)

where

p_i^* = vapor pressure of component i, at temperature

$P = \pi$ = total pressure

$\gamma = a_i/x_i$ = activity coefficient of component, i

p_i = partial pressure of component i.

Raoult's Law: $p_i = x_i\, p_i^* = y_1 P$

For “maximum” boiling binary azeotropes, the partial pressures will be less than predicted by Raoult's Law and the activity coefficients will be less than 1.0.

In reference to distillation conditions, the azeotrope represents a point in the system where the relative volatilities reverse. This applies to either type of azeotrope but the direction of reversal is opposite. For example in Figure 10-20 b the lower portion of the x-y diagram shows that $y_i > x_i$, while at the upper part, the $y_i < x_i$. In any distillation, without addition of an azeotrope “breaker” or solvent to change the system characteristics, if a feed of composition 30% x_i were used, the column could only yield a bottom product, which approaches pure x_2

and an azeotrope composition of about 65% x_i and 35% x_2 at the top. The situation would be changed only to the extent of recognizing that if the feed came in above the azeotropic point, the bottoms product would be the azeotrope composition.

Successful azeotropic distillation is largely dependent on the correct choice of entrainer (i.e. solvent). Entrainers are selected on the basis of making a seperation distillation possible. Stichlmair and Fair [263] have presented the criteria for entrainer selection, which are illustrated in Table 10-12 and Figure 10-25a. The entrainer, selected such that the constituents *a* and *b* of the azeotropic mixture to be separated are both origins or termini of distillation lines, and such systems always have distillation borders. The entrainer must also permit the recovery of products *a* and *b* at the top or bottom of distillation columns.

Table 10-12 Criteria for Entrainer Selection for Processes With Distillation Borders. See Process in Figure 10-23.
Source: Stichlmair, J.G., and J. R. Fair, Distillation: Principles and Practice, Wiley-VCH, 1998.

Entrainer for the separation of a mixture with a minimum azeotrope
• Low boiler (lower than the minimum azeotrope)
• Medium boiler, which forms a new minimum azeotrope with the low boiling constituent of the given mixture.
• High boiler, which forms new minimum azeotropes with both constituents of the given mixture. At least one of them has to boil lower than the azeotrope of the given mixture.
Entrainer for the separation of a mixture with a maximum azeotrope
• High boiler (higher than the maximum azeotrope).
• Medium boiler, which forms a new maximum azeotrope with the high boiling constituent of the given mixture.

- Low boiler, which forms new maximum azeotropes with both constituents of the given mixture. At least one of them has to boil higher than the azeotrope of the given mixture.

Figure 10-23

Process flow diagram for the separation of a binary mixture with a minimum azeotrope by using a low boiling entrainer. The feed is rich in a. Components a and b are obtained as pure bottoms products from columns C-1 and C-2 respectively.

Stichlmair and Fair [263] have presented a generalized process for separating out the higher boiling components, *b*. In this instance, the sequence of the separations differs from that in Figure 10-23. Both modifications of the generalized process are characterized by two separations performed on the convex side and one separation performed on the concave side of the border distillation line. Using the criteria selection of Table 10-12, a minimum azeotrope can be separated by an intermediate boiling or a high boiling entrainer instead of a low boiler, if additional azeotropes are formed. Smith [65] discusses azeotropic distillation in detail; Widalgo and Seider [264] have presented an excellent review of azeotropic distillation. References 153-157, 171, and 172 also describe azeotropic design techniques.

Generally, all activity coefficient models are capable of representing phase diagrams of the type shown in Figures 10-10c and 10-10d, but not always to same level of accuracy. It is essential therefore, to plot

the x-y diagram for each binary pair in the (multicomponent) mixture before using the VLE model in design calculations. If the predicted diagrams do not provide satisfactory agreement with experimental data, other models should be considered. Otherwise, a poor design could result.

Commencing with the equality of the fugacities of the species in the vapor and liquid mixtures in Equation 10-12:

(10-12)

and substituting the expressions for the mixtures fugacities in terms of the mole fractions, activity coefficients, and fugacity coefficients in Equations 10-13 and 10-14:

(10-87)

Consider an ideal binary mixture , at low pressure , where is the vapor pressure of species i. Substituting in Equation 10-87,

(10-88)

and

(10-89)

Adding Equations (10-88) and (10-89):

(10-90)

(10-91)

Equation 10-91 shows the relationship between the total pressure, P and the mole fraction of the more volatile species, which is characteristic of Raoult's law as shown in Figure 10-24a and Figure 10-24b for a benzene-toluene mixture. However, the mixture exhibits a positive deviation from ideality or Raoult's law, so Equation 10-90 becomes:

(10-92)

Figure 10-24a Boiling point diagram for benzene-toluene mixture, total pressure = 760 mm Hg.

Figure 10-24b VLE diagram for the mixture benzene and toluene at 760 mm Hg total pressure.

Therefore, if and are ≥ 1 for all compositions under isothermal boiling, then it is common for the bubble and dew point curves to reach a maximum at the same composition, that is the azeotropic point. This situation is illustrated in Figure 10-25a and Figure 10-25b for an ethyl acetate-ethanol mixture. Figure 10-25a shows the bubble and dew point curves on a T-x-y diagram at 1 atmosphere pressure (101.3 kPa). Note the minimum boiling azeotrope at 71.8°C, where $x_1 = y_1 = 0.54$. Feed streams having lower mole fractions cannot be purified beyond 0.54 in a distillation column, and streams having higher mole fractions produce distillate mole fractions that are bounded by 0.54 and 1.0. Consequently, the azeotropic composition is commonly referred to as a distillation boundary.

Figure 10-25a Phase diagram for the mixture of ethyl acetate and ethanol at 1 atm.

Figure 10-25b VLE diagram for the mixture ethyl acetate and ethanol at 1 atm.

Similarly, when the mixture exhibits a negative deviation from ideality or Raoult's law, and are , and both the bubble and dew point curves drop below the straight line that represents an ideal mixture. Figure 10-26a and Figure 10-26b respectively show the phase and VLE diagrams of a water-formic acid binary mixture at 1 atmosphere pressure (101.3 kPa). The T-x-y diagram shows the boiling point curving upwards above ideality, and consequently the system is said to have the maximum boiling azeotrope at 107.6°C, where $x_1 = y_1 = 0.41$. In this instance, feed streams having lower mole fractions cannot be purified beyond 0.41 water in the bottoms product of a column, and feed streams having higher mole fractions have a lower bound of 0.41 water in the mole fraction of the bottoms product. The azeotropes in Figure 10-25a and Figure 10-26a occur where , or equivalently, where $x_1 = y_1$. Therefore, azeotropic mixtures produce x-y curves that cross the 45° line at the azeotropic composition. The slope of the curve at the crossing point is <1.0 for minimum-boiling azeotropes and >1.0 for maximum-boiling azeotropes, as shown in Figure 10-25a and Figure 10-26a respectively.

Figure 10-26a Phase diagram for the mixture of water and formic acid at 101.3 kPa.

Figure 10-26b VLE diagram for the mixture of water and formic acid at 101.3 kPa.

At homogeneous azeotrope, $x_i = y_i$, i = 1, ….C, the equilibrium constant, K_i for species becomes unity. The equilibrium constant for species i in Equation 10-27 becomes:

(10-93)

where the degree of nonideality is expressed by the deviations from unity of the activity coefficients, , for the liquid phase, and the fugacity coefficients, for the vapor phase. At low pressure, and and Equation 10-93 reduces to:

(10-94)

Since the equilibrium constants for all of the species are unity at an azeotrope point, a simple distillation approaches this point, after which no further separation can occur. For this reason, an azeotrope is often referred to as a stationary, or fixed pinched point. For a minimum-boiling azeotrope where the deviations from Raoult's law are sufficiently large ,

phase splitting may occur and a minimum-boiling heterogeneous azeotrope may occur with a vapor phase in equilibrium with two liquid phases. A heterogeneous azeotrope occurs when the vapor-liquid envelope overlaps with the liquid-liquid envelope as shown in Figure 10-14. The author has developed phase diagrams of many binary systems from available literature using Microsoft Excel. These diagrams are shown in Appendix M and can be downloaded from the Elsevier companion website.

Example 10-4: Raoult's Law, Total Pressure Calculations for Multicomponent Mixture

A hydrocarbon liquid is a mixture at 55°F of the following components:

Component	Mole %
Isobutane	41.5
Pentane	46.5
n-Hexane	12.0
	100.0

A vaporizer is to heat the mixture to 190°F at 110 psia; determine the total pressure of the hydrocarbon liquid mixture.

Data from vapor pressure charts such as [48]:

Component	Vapor pressure, psia at 190°F
Isobutane	235.0
Pentane	65.0

n-Hexane	26.0

Specific gravity of pure liquid at 55°F [79]:

Component	Specific gravity, SpGr.
Isobutane	0.575
Pentane	0.638
n-Hexane	0.678
Moles in original liquid. Basis 100 gallons liquid	

Assume Raoult's Law:

Component	Mols of component
Isobutane	
Pentane	= 46.5 (8.33×0.638)/MW = 247.12/72.146 =
n-Hexane	= 12 (8.33×0.678)/MW = 67.77/86.172 =
Total	

Component	Mole fraction in the liquid phase	

isobutane	x_1 = 3.42/7.631	0.448
Pentane	= x_2 = 3.425/7.63	0.449
n-hexane	= x_3 = 0.786/7.631	0.103
Total		1.000

Mol fraction in vapor phase at 190°F. Raoult's Law:

(10-36)

(10-95)

y_1 = 0.448 (235)/[(0.449) (65) + (0.448) (235) + (0.103) (26)]
= 105.28/[29.185 + 105.28+ 2.678] = 105.28/[137.143]
= 0.767
y_2 = 0.449 (65)/137.143 = 0.212
y_3 = 0.103 (26)/137.143 = 0.0195
Σy_i = 0.998 (not rounded)
(i.e. the total pressure = sum of the partial pressures).
Because, P_{total} = (0.448) (235) + 0.449 (65) + 0.103 (26) = 137.14 psia

This is greater than the selected pressure of 110 psia. Therefore, for a binary mixture the results can be obtained without a trial-and-error

solution. But for the case of mixtures of three or more components, the trial-and-error assumption of the temperature for the vapor pressure will require a new temperature, redetermination of the component's vapor pressure, and repetition of the process until a closer match with the pressure is obtained.

Example 10-5: Determination of the Composition At An Azeotropic Temperature

The ethyl acetate (1) –ethyl alcohol (2) azeotrope has a boiling point of 71.8°C at a pressure of 101.325 kN/m^2 and a composition of 53.9 mole % ethyl acetate.

(a) Determine the composition of the azeotrope at 95°C from the table below.

(b) Comment briefly upon the implications of changing the pressure of an azeotropic distillation operation.

Temperature °C	**Vapor Pressure**	**kN/m^2**
	Ethyl Acetate	**Ethyl Alcohol**
70	79.03	72.55
80	110.75	108.32
90	151.38	158.27
100	195.11	217.49

The van Laar equations are:

(10-96)

or, by rearrangement,

(10-97)

Solution

(a) The mole fraction in the liquid phase of ethyl acetate = x_1

The mole fraction in the vapor phase of ethyl acetate = y_1

The mole fraction in the liquid phase of ethyl alcohol = x_2

The mole fraction in the vapor phase of ethyl alcohol = y_2

At equilibrium and azeotropic conditions the following relationships hold:

(10-12)

where

= fugacity of component i in the vapor phase

= fugacity of component i in the liquid phase.

For a perfect gas and non ideal solution

In the vapor phase, the fugacity:

(10-98)

where π = the total pressure.

For nonideal liquid solution:

In the liquid phase, the fugacity:

(10-99)

where

γ_1 = activity coefficient of ethyl acetate. It is a measure of the deviation from Raoult's law. It is greater than unity for positive deviations and less than unity for negative deviation.

= vapor pressure of ethyl acetate.

Activity coefficients are defined for each component of a mixture as:

(10-100)

and

(10-101)

At azeotropic composition the values $x_1 = y_1$ and $x_2 = y_2$; so the activity coefficients become:

(10-102)

and

(10-103)

Dividing Eq. 10-102 by Eq. 10-103 gives:

(10-104)

At the boiling point of 71.8°C, the vapor pressures of ethyl acetate, and ethyl alcohol, are by interpolation:

Ethyl acetate component	Ethyl alcohol component
= 85 kN/m^2	= 79 kN/m^2

The activity coefficients at the boiling temperature of 71.8°C are:

and

The level of ethyl acetate in the mixture is 53.9 mole % and that of ethyl alcohol is 46.1 mole %.

The van Laar coefficients A and B are:

= 0.859

= 0.829

Using the values of A and B, the activity coefficients γ_1 and γ_2 are computed for various values of x_1 and x_2 from the equations as shown below. The values of γ_1 and γ_2 are shown in the subsequent table.

x_1	x_2	γ_1	γ_2	
0.4	0.6	1.3505	1.1484	1.176
0.6	0.4	1.1407	1.3593	0.839
0.8	0.2	1.0329	1.7127	0.603

At the azeotrope temperature of 95°C, the values of and are determined by interpolation as follows:

and

and

The corresponding liquid phase mole fraction x_1 is calculated by interpolation as follows:

At 95°C, $x_1 = 0.454$

(b) Changing the pressure infers (by Charles' Law) that the temperature, and therefore the boiling point will change directly. The composition will also change, so that an azeotrope may be avoided.

10.13. Bubble Point of Liquid Mixture

An equilibrium between a liquid and a vapor implies that the liquid is on the point of boiling, is producing bubbles of vapor within it, and the vapor is on the point of condensing or forming a dew. These conditions are referred to as saturated liquid and saturated vapor. At the prevailing total pressure, the liquid is at its bubble point temperature and the vapor at its dew point temperature. The amount of the vapor formed is assumed to be so small as to have no effect on the liquid composition. For a binary system:

(10-105)

and

(10-106)

In the majority of cases involving multicomponent mixtures, no direct solution to Equation 10 –105 exists. The basic definition of the equilibrium coefficient is:

(10-107)

For a stable system, the total mole fraction is unity, *i.e.*

(10-108)

Equation 10-108 is used to determine the bubble point temperature and pressure. In using it, the temperature or pressure is fixed, while the other parameter is varied until the criterion for a stable system is identified. A combination of temperature and pressure is altered, if the summation of the calculated vapor composition differs from unity. There is no direct method that will allow a reasonable estimation of the amount of change required. However, Dodge [265], Hines and Maddox [266] have provided techniques for reducing the number of trials required.

10.13.1. Dew point calculations

The dew point of a vapor is that combination of temperature and pressure at which the first drop of the liquid condenses. The dew point criterion is:

(10-109)

For binary mixtures of components 1 and 2, then:

(10-110)

and

(10-111)

(10-111)

These two equations can be used to calculate the composition of a binary vapor that will begin to condense at a given temperature and pressure.

Example 10-6: Dewpoint Temperature Calculations For a Multicomponent Mixture

A multicomponent hydrocarbon vapor mixture is of composition:

Component	Mol %	K
Methane	27.52	7.88
Ethane	16.34	2.77
Propane	29.18	1.18
Isobutane	5.37	0.61
Normal butane	17.18	0.48
Isopentane	1.72	0.264
Normal pentane	2.18	0.225
Hexane	0.47	0.102
Heptane	0.04	0.048

Operating temperature and pressure are 178°C and 400 psia. Determine the dew point temperature of the mixture.

Solution

The computer program PROG101 calculates the dew and bubble points of any multicomponent hydrocarbon mixture based on user supplied K-values (see Figure 10-12aFigure 10-12bFigure 10-12c and Figure 10-12d). The program will handle feed streams containing up

to 15 components. The feed entries may be expressed in moles, mole fraction or mole %. The dew point should be 1.0 or very close to 1.0, depending on the acceptable tolerance. If the sum obtained is not equal to 1.0 (within the user tolerance), a message is shown instructing the user to assume a new temperature. The iteration is repeated using the corresponding to the new temperature selected. The temperature supplied does not affect the computation, but is printed and stored as a reminder to the user of which temperature (and corresponding K-values) was used for the last iteration. Program PROG101 can be run from its executable file (PROG101.EXE) generated from the latest Absoft Pro Fortran version 10 compiler using the Microsoft Run time Windows Environment (MRWE) application. The user only needs to supply the data file (e.g. DATA101.DAT) and run the executable PROG101.exe to generate the results. Both the source and executable codes can be downloaded from the Elsevier companion website. Table 10-13 shows the input data and results instructing the user that the dew point temperature should be increased. Appendix I illustrates the steps followed in creating and running the program.

Table 10-13 Input Data and Computer Results for Dew and Bubble Points Calculations.

Data name: Data101. Dat	
Dew	
178.0	
400.0	
9	
27.52	7.88
16.34	2.88

29.18	1.18
5.37	0.61
17.18	0.48
1.72	0.264
2.18	0.225
0.47	0.102
0.04	0.048
Dew and bubble points calculations at 178.0°F and 400.0 psia	
Dew point calculation	

Temperature °F: 178.00		
Number of components in the feed stream: 9		
Number of components	**Feed composition (Mole %)**	**K-Value**
1	27.5200	7.8800
2	16.3400	2.7700
3	29.1800	1.1800
4	5.3700	0.6100
5	17.1800	0.4800
6	1.7200	0.2640
7	2.1800	0.2250
8	0.4700	0.1020
9	0.0400	0.0480
Total composition:	100.0000	
Number of components	Mole fraction	X = Y/K
1	0.2752	0.0349
2	0.1634	0.0590
3	0.2918	0.2473
4	0.0537	0.0880
5	0.1718	0.3579
6	0.0172	0.0652

7	0.0218	0.0969
8	0.0047	0.0461
9	0.0004	0.0083
Total dew point:		1.0036
Raise the dewpoint temperature of: 178.00°F		

Example 10-7: Bubble-point temperature Calculations For Multicomponent Mixture

At 200 psia, calculate the bubble point temperature of the following:

Component	Mol %	K @ 260°F	K @ 235°F	K @ 237°F
	18.2	1.92	1.62	1.65
	23.8	1.58	1.35	1.35
	33.7	0.93	0.76	0.77
	12.1	0.81	0.64	0.64
	12.2	0.42	0.315	0.32

Solution

The computer program PROG101 determines the bubble points of the hydrocarbon mixture. As with the dew point, the bubble point is computed until the sum = 1.0, or close to this value, obtained depending on the accuracy required. If the sum is not equal to 1.0 (within tolerance), the iteration is repeated using the corresponding to the new temperature. Table 10-14Table 10-15 and Table 10-16 give input data and results for the bubble points of the hydrocarbon mixture at 260°F, 235°F and 237°F respectively. By interpolation, the bubble temperature is 237.3°F at 200 psia pressure.

Table 10-14 Input Data and Computer Results for Dew and Bubble Points

Table 10-14 Input Data and Computer Results for Dew and Bubble Points Calculations.

Data name: Data101. Dat	
Bubble	
260	
200	
5	
18.2	1.92
23.8	1.58
33.7	0.93
12.1	0.81
12.2	0.42
Dew and bubble points calculations at 260.0°F and 200.0 psia	
Bubble point calculation	

Temperature of: 260.00		
Number of components in the feed stream: 5		
Number of components	**Feed composition (mole %)**	**K-Value**
1	18.2000	1.9200
2	23.8000	1.5800
3	33.7000	0.9300
4	12.1000	0.8100
5	12.2000	0.4200
Total composition:	100.0000	
Number of components	Mole fraction	Y=KX
1	0.1820	0.3494
2	0.2380	0.3760
3	0.3370	0.3134
4	0.1210	0.0980
5	0.1220	0.0512

Total bubble point:		1.1881
Lower the bubble point temperature of: 260.00°F		

Table 10-15 Input Data and Computer Results for Dew and Bubble Points Calculations.

Data name: Data101. Dat		
Bubble		
235		
200		
5		
18.2	1.62	
23.8	1.35	
33.7	0.76	
12.1	0.64	
12.2	0.315	
Dew and bubble points calculations at 235.0°F and 200.0 psia		
Bubble point calculation		
Temperature of: 235.00		
Number of components in the feed stream: 5		
Number of components	**Feed composition (Mole %)**	**K-Value**
1	18.2000	1.6200
2	23.8000	1.3500
3	33.7000	0.7600
4	12.1000	0.6400
5	12.2000	0.3150
Total composition:	100.0000	
Number of components	Mole fraction	Y=KX
1	0.1820	0.2948

1	[illegible]	[illegible]
2	0.2380	0.3213
3	0.3370	0.2561
4	0.1210	0.0774
5	0.1220	0.0384
Total bubble point:		0.9881
Raise the bubble point temperature of: 235.00°F		

Table 10-16 Input Data and Computer Results for Dew and Boiling Points Calculations.

Data name: Data101. Dat		
Bubble		
237		
200		
5		
18.2	1.65	
23.8	1.35	
33.7	0.77	
12.1	0.64	
12.2	0.32	
Dew and bubble points calculations at 237.0°F and 200.0 psia		
Bubble point calculation		
Temperature of: 237.00		
Number of components in the feed stream: 5		
Number of components	**Feed composition (Mole %)**	**K-Value**
1	18.2000	1.6500
2	23.8000	1.3500
3	33.7000	0.7700
4	12.1000	0.6400

4	12.1000	0.6400
5	12.2000	0.3200
Total composition:	100.0000	
Number of components	Mole fraction	Y=KX
1	0.1820	0.3003
2	0.2380	0.3213
3	0.3370	0.2595
4	0.1210	0.0774
5	0.1220	0.0390
Total bubble point:		0.9976
Raise the bubble point temperature of: 237.00°F		

10.14. Equilibrium Flash Computations

Flash vaporization calculations involving multicomponent mixtures are necessary for numerous processes, often to determine the condition of the feed to a fractionating column, or to determine the flow of vapor from reboiler or condenser.

Such calculations often involve trial-and-error solutions, which are time-consuming, tedious and subject to error if performed manually. They are better carried out with computer-aided process design and simulation programs, such as UniSim® (Honeywell) Design, Aspen Plus, HYSYS, ChemCad, and Pro II. This author [267] has developed an improved iterative convergence method first suggested by Oliver [268], and later modified by Kostecke [269], for isothermal equilibrium flash computations. Multicomponent flash (isothermal and adiabatic) computations are now incorporated as part of an overall process simulation and equipment design package. However, single-stage flash fractionation processes are also employed to separate light components in a feed, and as a preliminary step before a multicomponent fractionation column, *e.g.* crude oil distillation. Table 10-17 lists ways of producing twophase mixtures from a single phase

under appropriate conditions. The last process in the table typifies the well-head separation that takes place in an oil field. It is termed a flash process because the vapor forms due to the rapid drop in the pressure.

Table 10-17 Ways to Produce TwoPhase Mixture.

Initial	Action to Produce TwoPhase Mixture
Gas	Cool, possibly after initial compression
Gas	Expand through a valve or an engine
Liquid	Heat to achieve partial vaporization
Liquid	Reduce pressure through a valve, if close to saturation

10.14.1. Fundamentals

To carry out an appropriate flash calculation, the pressure, P and the temperature, T, must be known. If the values of P and T in the separating vessel are fixed, the value of P must not be so high that the two phases cannot exist at any value of T. Nor must T lie outside the bubble point and dew point range corresponding to P. For a valid twophase equilibrium calculation, the following relationship must be satisfied:

(10-112)

where

T_{bp} = the bubble point temperature

T_s = the specified temperature

T_{dp} = the dew point temperature

The existence of a valid twophase flash point can be verified with the design equations for the bubble point, dew point, and equilibrium data

calculated at the specified pressure and temperature.

10.14.2. Calculation Of Bubble Point And Dew Point

The design equations for the bubble point and dew point are:

Bubble point

(10-113)

Dew point

(10-114)

The calculation method is as follows (at constant pressure):

1. Assume a temperature, T.
2. Calculate K –values.
3. Calculate the sum of the right side of Equation 10–113 for a bubble point calculation. If < 1.0, increase the temperature. If > 1.0, decrease the temperature. Repeat steps 2 and 3 until convergence is attained.
4. For a dew point calculation, obtain the right side of Equation 10-114. If < 1.0, decrease the temperature. If > 1.0, increase the temperature. Repeat steps 2 and 4 until a convergence is reached.

Table 10-18 illustrates the phase condition using the light-vapor data associated with the specified pressure and temperature. From Table 10-18, the conditions for a valid twophase equilibrium flash are:

(10-115)

and

(10-116)

Table 10-18 Equilibrium Flash Criteria.

Subcooled liquid	<1	>1
Bubble point	=1	>1
Twophase condition	>1	>1
Dew point	>1	=1
Superheated vapor	>1	<1

Feed analyses of component concentrations are generally unavailable for complex hydrocarbon mixtures with a final boiling point of >38°C, unless the feed is subdivided into pseudo-components (narrow boiling fractions). This subdivision enables the mole fraction and equilibrium constant, K, to be estimated, and consequently, a flash calculation can be carried out for the mixture. A source of K values for light hydrocarbon systems is the DePriester [80] charts, which are shown in Figure 10-12aFigure 10-12bFigure 10-12c and Figure 10-12d. These give the K values over a wide range of temperature and pressure. Because the charts are nomographs, a straight line connecting the temperature and pressure for which K values are required will intersect curves for each compound at its K value. The

values have been obtained by calculating fugacities from an equation of state. Use of these charts for exact determination of K values requires a trial-and-error approach. Hadden and Grayson [270] have presented correlations for hydrocarbon vapor-liquid distribution ratios, and the charts from their work are shown in Figure 10-27 and Figure 10-28. These charts can be readily used for determining the vapor-liquid distribution coefficients. Figure 10-29 shows a continuous equilibrium flash fractionation process.

Figure 10-27

Vapor-liquid equilibrium constants, −260 to + 100°F

(source: Hadden and Grayson [270]).

Figure 10-28

Vapor-liquid equilibrium constants, 40 to 800° F

(source: Hadden and Grayson [270]).

Figure 10-29

Continuous equilibrium flash fractionation.

The Equations

The following equations are used for the multicomponent equilibrium flash calculations.

(10-117)

(10-118)

(10-119)

where

R = the liquid-vapor (L/V) ratio. The new L/V ratio for each iterative calculation, R′, is determined from:

(10-120)

The constants E and F are determined by:

(10-121)

and

(10-122)

(10-122)

The Algorithm

The computation then determines whether . If is $\not<$ 0.001, R is set equal to R′ and the calculations of Equations (10-119)(10-120)(10-121) and (10-122) are repeated until .

Once , the calculation proceeds to find the total moles of component i in both the liquid and vapor phases, that is L_i and V_i. These are given by:

(10-123)

and

(10-124)

The mole fractions of components in the feed, the liquid and vapor phases are evaluated as follows:

(10-125)

(10-126)

and

(10-127)

where

(10-128)

where

E, F = constants

K_i = equilibrium flash constant for each component in the feed stream

L_i = total moles of component i in the liquid phase

L = total moles of liquid at equilibrium conditions

M_i = total moles of component i the feed stream

n_i = mole fraction of component i in the feed stream

R = liquid-vapor (L/V) ratio

R' = new L/V ratio for each iteration calculation

V = total moles of vapor at equilibrium conditions

V_i = total moles of component i in the vapor phase

x_i = mole fraction of component i in the liquid phase

y_i = mole fraction of component i in the vapor phase

The method described here is based on the vapor-liquid equilibrium relationships given in Engineering Data books available from the Gas Processors Suppliers Association [79].

The Program

A Fortran program named PROG102 has been developed for flash calculation; it is based upon the vapor-liquid equilibrium relationships

given in Engineering Data books available from the Gas Processors Suppliers Association, Tulsa and other literature. It will handle calculations with feed streams containing up to 15 components. In addition, the calculation will check the feed composition at flash conditions for dew and bubble points (i.e. whether the feed is either all vapor or all liquid). These checks are performed before the flash calculations are started. If the feed is above the dew point or below the bubble point, an appropriate message is displayed on the screen. A default value of R (L/V ratio) = 1 is incorporated in the program to start the iterative process. The program is not applicable to an adiabatic flash process, when the feed stream is at pressure higher than the flash pressure, and the heat of vaporization is provided by the enthalpy of the feed. In this situation, the flash temperature will be unknown and must be obtained by trial and error. A temperature must be found at which both the material and energy balances are satisfied.

PROG102 has been developed using the Absoft v.10 Fortran compiler with Microsoft Run time Windows Environment (MRWE) application. The user need only create the data source file (e.g. DATA102.DAT) and run the executable PROG102.exe to generate the results. Both the source and executable codes can be downloaded from the Elsevier companion website.

Example 10-8: Determination of the Amount of Liquid and Vapor in A MultiComponent Mixture

Determine the amount of liquid and vapor in a nine-component still product accumulator stream that is compressed to a pressure of 370 psia and cooled to 90°F.

Process Data

Number	**Component**	**Feed mols/hr**	**Equilibrium constant, K**

1		2,752	7.2
2		1,634	1.65
3		2,918	0.54
4		537	0.25
5		1,718	0.185
6		172	0.088
7		218	0.069
8		47	0.028
9		4	0.00078

Solution

The computer program PROG102 calculates the liquid and vapor streams from the equilibrium constants and molar flow rates of the components at the supplied temperature and pressure. Table 10-19 shows the input data and results from PROG102.

Table 10-19 Data Input and Computer Results of Multicomponent Equilibrium Flash Calculations

Data name: Data102.Dat	
90	
370	
9	
2752.0	7.2
1634.0	1.65
2918.0	0.54
537.0	0.25
1718.0	0.185
172.0	0.088

172.0	0.088
218.0	0.069
47.0	0.028
4.0	0.00078

Multicomponent equilibrium flash calculation							
Multicomponent equilibrium flash calculation at 90.0°F and 370.0 psia							
Component number	**K-Value**	**Feed**		**Liquid**		**Vapor**	
		Mols/h.	**Mol. frac.**	**Mols/h.**	**Mol. frac.**	**Mols/h.**	**Mol. frac**
1	7.200	2752.000	0.275	388.353	0.072	2363.647	0.516
2	1.650	1634.000	0.163	682.316	0.126	951.684	0.208
3	0.540	2918.000	0.292	2003.467	0.370	914.533	0.200
4	0.250	537.000	0.054	443.314	0.082	93.686	0.020
5	0.185	1718.000	0.172	1485.664	0.274	232.336	0.051
6	0.088	172.000	0.017	160.091	0.030	11.909	0.003
7	0.069	218.000	0.022	205.985	0.038	12.015	0.003
8	0.028	47.000	0.005	45.913	0.008	1.087	0.000
9	0.001	4.000	0.000	3.997	0.001	0.003	0.000
TOTAL		10000.000	1.000	5419.101	1.000	4580.899	1.000

Example 10-9: Determination Of The Flow Rates of Liquid and Vapor in a Gas Liquefaction Plant

A feed stream of natural gas is flashed at a pressure of 600 psia and 20°F in a gas liquefaction plant. Determine the flow rates of the liquid and vapor streams of a feed flow rate of 1,000 mols/hr.

Process Data

Number	**Component**	**Feed Moles fraction**	**Equilibrium constant, K**
1		0.0112	0.90

1		0.0112	0.90
2		0.8957	2.70
3		0.0526	0.38
4		0.0197	0.098
5		0.0068	0.038
6		0.0047	0.024
7		0.0038	0.0075
8		0.0031	0.0019
9	and heavier	0.0024	0.0007

Solution

Table 10-20 shows the input data and results of an isothermal equilibrium flash calculation at 20°F and 600 psia.

Table 10-20 Data Input and Computer Results of Multicomponent Equilibrium Flash Calculations

Data name: Data102.Dat	
20.0	
600.0	
9	
11.2	0.90
895.7	2.70
52.6	0.38
19.7	0.098
6.8	0.038
4.7	0.024
3.8	0.007
3.1	0.0019

3.1	0.0015
2.4	0.0007

Multicomponent equilibrium flash calculation							
Multicomponent equilibrium flash calculation at 20.0°F and 600.0 psia							
Component number	**K-Value**	**Feed**		**Liquid**		**Vapor**	
		Mols/h.	**Mol. frac.**	**Mols/h.**	**Mol. frac.**	**Mols/h.**	**Mol. frac**
1	0.900	11.200	0.011	0.506	0.012	10.694	0.011
2	2.700	895.700	0.896	13.903	0.340	881.797	0.919
3	0.380	52.700	0.053	5.299	0.130	47.301	0.049
4	0.098	19.700	0.020	5.966	0.146	13.734	0.014
5	0.038	6.800	0.007	3.593	0.088	3.207	0.003
6	0.024	4.700	0.005	3.006	0.074	1.694	0.002
7	0.007	3.800	0.004	3.231	0.079	0.569	0.001
8	0.002	3.100	0.003	2.968	0.073	0.132	0.000
9	0.001	2.400	0.002	2.361	0.085	0.039	0.000
TOTAL		1000.000	1.000	40.831	1.000	959.169	1.000

10.15. Degrees of Freedom

The design of separation systems involves many parameters, and the number of variables that can be arbitrarily specified is known as the degrees of freedom. This is determined by subtracting the number of thermodynamic equilibrium equations from the number of variables. For nonreacting systems, the Gibbs phase rule is defined by:

(10-129)

where

F = degrees of freedom (i.e. the number of independent properties that have to be specified to determine all the intensive properties of each phase of the system of interest).

C = number of components in the system. For circumstances that involve chemical reactions, C is not identical to the number of chemical compounds in the system but is equal to the number of chemical compounds less the number of independent-reaction and other equilibrium relationships among these compounds.

P = number of phases that can exist in the system. A phase is a homogeneous quantity of material *e.g.* gas, pure liquid, a solution, or a homogeneous solid.

Variables of the kind with which the phase rule is concerned are termed phase-rule variables, and they are inherent properties of the system, *i.e.* properties that do not depend on the quantity of material present.

For example, in a binary system such as methanol-water, having liquid and vapor phases,

$C = 2$, $P = 2$ and:

$F = 2 - 2 + 2 = 2$

When temperature and pressure are set, all the degrees of freedom are used, and at equilibrium all compositions are determined from experiment. Alternatively, the pressure and components' mole fractions (methanol and water) and determine the temperature and the other mole fractions. The phase rule refers to inherent variables where temperature, pressure, specific (per unit mass) values (i.e. specific volume) or mole fractions do not depend on the total amount of material present. The extensive variables *i.e.* number of moles, flow rate and volume depend on the amount of material and are not included in the degree of freedom.

Another example is the ideal gas equation , where if the phase rule

is applied for a single phase and for a pure gas, P = 1, C = 1, then

F = 1 – 1 + 2 = 2 variables to be specified. Since the phase rule is concerned with intensive properties only, the following phase-rule variables in the ideal gas law:

By specifying two intensive variables (F = 2), the third variable can be determined.

When a set of independent equations representing the material balances for a problem is prepared, the number of variables whose values are unknown is counted. If more variables whose values are unknown exist than independent equations, then an infinite number of solutions exist for a material balance, and such problems are referred to as *underspecified (underdetermined)*. Either values of additional variables must be found to make up the deficit or the problem must be posed as an optimization problem. Alternatively, if fewer values of the variables whose values are unknown exist than the independent equations, the problem is *over specified (over determined)* and no solution exists to the problem, showing that the equations are inconsistent. In this instance, the problem might be posed as an optimization problem to minimize the sum of the squares of the deviations of the equations from zero (or their right-hand constants).

The difference between the number of variables whose values are unknown and the number of independent equations is termed the number of degrees of freedom. If the degrees of freedom are positive, such as 2, two additional independent equations or specifications of variables must be sought to provide a unique solution to the material balance problem. Conversely, if the degrees of freedom are negative *e.g.* −1, there are many equations or not enough variables in the

problem. *Zero degrees of freedom means that the material balances problem is properly specified*, and the equations can be solved for the variables whose values are unknown. In this instance, the number of degrees of freedom is the number of variables in a set of independent equations to which values must be assigned so that the equations can be solved. This can be expressed by:

(10-130)

where

N_d = number of degrees of freedom

N_v = number of variables

N_r = number of equations

10.16. UniSim (Honeywell) Software

UniSim® Design R360.1 simulation software is an intuitive and interactive process modeling package that allows engineers to create steady-state and dynamic models for plant design, performance monitoring, troubleshooting, operational improvement, business planning and asset management. Further, it provides users with Honeywell's domain expertise in process simulation and operator training. Using industry-specific unit operating models and powerful tools to optimize operating parameters for feedstock changes, UniSim Design allows the steady-state simulation of oil and gas, refining, chemical and petrochemical processes.

In this book, UniSim Design simulation software is employed to verify some design calculations. Appendix-K shows the steps in creating a simulation run using UniSim Design R360.1.

Example 10-10: Application of UniSim Design Software for Example 10-8

Example 10-8 is repeated using UniSim Design R360.1 software (Example 10-10.usc) to determine the amount of liquid and vapor in a nine-component still product accumulator stream that is compressed to 370 psia and cooled to 90°F. Figure 10-30 shows a screen shot of UniSim Design 360.1.

Figure 10-30

a snapshot of the Process flow diagram of a flash drum

(courtesy of Honeywell UniSim® Design).

Table 10-21 shows the simulation results from Honeywell UniSim when the Peng Robinson equation for vapor liquid equilibrium (VLE) is used. The results for the total vapor and liquid flow rates give good agreement with the results in Table 10-19.

Table 10-21 Results of the simulation using UniSim® Design Software for Example 10-8.

10.17. Binary System Material Balance: Constant Molal Overflow Tray to Tray

Refer to Figure 10-5. (For an overall review, see Reference 173)

- Rectifying Section:

For any component in the mixture; using total condenser see Figure 10-5 and Figure 10-33.

(10-132)

(10-133)

• Operating Line Equation:

(10-134)

For total condenser: y (top plate) = x_D

• Stripping Section:

(10-135)

(10-136)

• Operating Line Equation:

(10-137)

10.17.1. Conditions of Operation (usually fixed)

1. Feed composition, and quantity.
2. Reflux Ratio (this is one of the initial unknowns).
3. Thermal condition of feed (e.g. liquid at boiling point, all vapor, subcooled liquid).
4. Degree, type or amount of fractionation or separation, including compositions of overhead or bottoms.
5. Column operating pressure or temperature of condensation of overhead (determined by temperature of cooling medium), including type of condensation, i.e., total or partial.
6. Constant molal overflow from stage to stage (theoretical) for simple ideal systems following Raoult's Law. Techniques that are more complicated apply for nonideal systems.

10.17.2. Flash Vaporization (see Figure 10-31)

At a total pressure, P, the temperature of flash must be between the dew point and the bubble point of a mixture [144 − 148]. Thus:

Figure 10-31

Schematic liquid flash tank. Note: Feed can be preheated to vaporize feed partially.

T (Bubble Point) < T (Flash) < T (Dew Point)

For binary mixtures [147]:

1. Set the temperature and pressure of the flash chamber.
2. Perform a material balance on a single component.

3. Knowing F, calculate the amounts and compositions of V and L.

$F_t = F + V_s$ = mols of feed plus mols of noncondensable gases

From Henry's Law:

(10-139)

$V_t = V + V_s$ = mols of vapor formed plus mols noncondensed gases

4.

(10-140)

(10-141)

(10-142)

After calculating V, calculate the x_i's and y_i's:

(10-143)

Then,

(10-144)

Calculate each y; after calculating the y_i's, calculate x_i's as follows:

(10-38)

so, $x_i = y_i/K_i$

(10-145)

Relation for K:

(10-38)

(10-146)

(10-147)

(10-148)

(10-149)

where

X_i = mols of a component i in vapor plus mols of same component in liquid, divided by total mols of feed (both liquid and vapor)
= total mol fraction, irrespective of whether component is in liquid or vapor

(10-150)

(10-151)

(10-152)

(10-153)

To calculate, V, L, y_i's, and x_i's:

1. Assume: V
2. Calculate: L = F − V
3. Calculate: L/V

4. Look up K_i's at temperature and total pressure of system
5. Substitute in:

(10-154)

6. If an equality is obtained from:
$V_{calc} = V_{assumed}$ the amount of vapor was satisfactory as assumed.

(10-155)

7.

(10-143)

and

(10-144)

8. Calculate each y_i as in (7) above, then the x_i's are determined:
$y_i = K_i x_i$
or, $x_i = y_i/K_i$
9.

(10-145)

where

n = total system pressure absolute.

P_i = vapor pressure of individual component at temperature,

abs.

X_i = X = mols of a component i, in vapor phase plus mols of same component in liquid divided by the total mols of feed (both liquid and vapor).

x_f = mol fraction of a component in feed.

x_f = mol fraction of any component in the feed, F_t where $X_f = Fx_f/F_t$ for all components in F; for the noncondensable gases, $x_f = V_s/F_t$.

F = mols of feed entering flash zone per unit time contains all components except noncondensable gases.

$F_t = F + V_s$

$V_t = V + V_s$ mols of vapor at a specific temperature and pressure, leaving flash zone per unit time.

V_s = mols of noncondensable gases entering with the feed, F, and leaving with the vapor, V, per unit time.

V = mols of vapor produced from F per unit time

F = V + L

L = mols of liquid at a specific temperature and pressure, from F, per unit time

i = specific individual component in mixture

K_i = equilibrium K values for a specific component at a specific temperature and pressure, from References 18, 65, 79, 99, 131, 235

T = temperature, abs

x_i = mol fraction of a specific component in liquid mixture as may be associated with feed, distillate, or bottoms, respectively

y_i = mol fraction of a specific component in vapor mixture as may be associated with the feed, distillate or bottoms, respectively

10. For the simplified case of a mixture free of noncondensable

gases, see Equation 10-142, where $X_{fi} = x_f$.

Example 10-11: Calculation of Bubble Point and Dew Point

From the hydrocarbon feed stock listed, calculate the bubble point and dew point of the mixture at 165 psia, and using K values as listed, which can be read from a chart in 3rd edition Perry's, Chemical Engineer's Handbook.

Feed Stock

Composition	Mol Fraction
C_2H_6	0.15
C_3H_8	0.15
n-C_4H_{10}	0.30
i-C_4H_{10}	0.25
n-C_4H_{12}	0.15
	1.00

Calculate the Bubble Point (Assume composition is liquid)

Composition	Mol Fraction	Assumed T = 90°F, K	Kx	Assume T = 100°F, K	Kx
C_2H_6	0.15	3.1	0.465	3.4	0.5100
C_3H_8	0.15	1.0	0.150	1.2	0.1800
n-C_4H_{10}	0.30	0.35	0.105	0.39	0.1170
i-C_4H_{10}	0.25	0.46	0.115	0.52	0.1300
n-C_5H_{12}	0.15	0.12	0.018	0.13	0.0195
	1.00		0.853		0.9565

					(Too low)
		Assume T = 105°F			
			K	Kx	
			3.45	0.5175	
			1.25	0.1875	
			0.41	0.1230	
			0.55	0.1375	
			0.15	0.0225	
				0.9880	

By interpolation:

X = 106.9°F (107°F)

So, T = 107°F Bubble point at 165 psia.

Calculation of Dew Point:

Composition in vapor	**Mol. Frac. in vapor**	**Assume, T = 160°F, K (from charts)**	**x = y/K**
C_2H_6	0.15	5.1	0.0294
C_3H_8	0.15	1.85	0.0811
n-C_4H_{10}	0.30	0.80	0.3750

i-C_4H_{10}	0.25	1.00	0.2500
n-C_5H_{12}	0.15	0.32	0.4688
	1.00		1.204 ≠ Σy/K = 1

Composition in vapor	Mol. Frac. in vapor	Assume, T = 180°F, K (from charts)	x = y/K
C_2H_6	0.15	5.95	0.0252
C_3H_8	0.15	2.25	0.0666
n-C_4H_{10}	0.30	0.98	0.3061
i-C_4H_{10}	0.25	1.34	0.1866
n-C_5H_{12}	0.15	0.41	0.3659
	1.00		0.9504 ≠ Σy/K = 1

Composition in vapor	Mol. Frac. in vapor	Assume, T = 175°F, K (from charts)	x = y/K
C_2H_6	0.15	5.60	0.0268
C_3H_8	0.15	2.20	0.0682
n-C_4H_{10}	0.30	0.91	0.3297
i-C_4H_{10}	0.25	1.20	0.2083
n-C_5H_{12}	0.15	0.39	0.3846
	1.00		1.0176 ≅ Σy/K = 1

Dew point is essentially 175°F at 165 psia.

Example 10-12: Calculation of Flashing Composition

If the mixture shown in Example 10-11 is flashed at a temperature midway between the bubble point and dew point, and at a pressure of 75 psia, calculate the amounts and compositions of the gas and liquid phases.

Referring to Example 10-11:

Composition	**Mol Fraction**
C_2H_6	0.15
C_3H_8	0.15
n-C_4H_{10}	0.30
i-C_4H_{10}	0.25
n-C_5H_{12}	0.15
	1.00

Must Calculate Bubble Point at 75 psia

Composition	**Mol Frac.**	**K @ 50°F**	**y = Kx**	**K @ 40°F**	**y = Kx**
C_2H_6	0.15	5.0	0.7500	4.5	0.6750
C_3H_8	0.15	1.2	0.1800	1.07	0.1605
n-C_4H_{10}	0.30	0.325	0.0975	0.28	0.0840
i-C_4H_{10}	0.25	0.48	0.1200	0.415	0.10380
n-C_5H_{12}	0.15	0.089	0.0134	0.074	0.0110
			1.1609		1.0343

Bubble point = 40°F (as close as K curves can be read)

Extrapolating =

Therefore, a close value of bubble point would be: 40° − 3° = 37°F

Calculate Dew Point at 75 psia

Composition	**Mol Frac.**	**K@ 70°F**	**Σx = y/K**	**K@ 100°F**	**Σx = y/K**	**K@ 130°F**	**Σx = y/K**
C_2H_6	0.15	5.9	0.0254	7.6	0.0197	9.5	0.0158
C_3H_8	0.15	1.55	0.0968	2.18	0.0688	3.0	0.0500
n-C_4H_{10}	0.30	0.45	0.6667	0.70	0.4286	1.06	0.2830
i-C_4H_{10}	0.25	0.58	0.4310	1.0	0.2500	1.45	0.1724
n-C_5H_{12}	0.15	0.13	1.1538	0.225	0.6667	0.37	0.4054
			2.3737		1.4338		0.9266
		(too low temp.)					

Refer to extrapolation curve, Figure 10-32.

Figure 10-32

Extrapolation curve for dew point for Example 10-12.

At $\Sigma x = y/K = 1.0$, dew point = 124°F

Flash this mixture at temperature midway between bubble point and dew point, or flash temperature =

Assume: F (feed) = 100

Pressure: = 75 psia; then tabulating the calculations:

Composition	**Feed Mol. Frac., x**	**FX**	**L/V**	**K@ 81°F**	
C_2H_6	0.15	15		6.6	0.3545
C_3H_8	0.15	15		1.78	1.3146
n-C_4H_{10}	0.30	30	2.34	0.54	4.3333
i-C_4H_{10}	0.25	25		0.77	3.0390
n-C_5H_{12}	0.15	15		0.16	14.6250

Vapor phase after flashing at 75 psia and 81°F = 30.3% of original feed.

Liquid phase = 100 − 30 = 70% of original feed

Composition of Vapor

Composition	**Mol Fraction y**
C_2H_6	11.074/30.33 = 0.3651
C_3H_8	6.481/30.33 = 0.2137
n-C_4H_{10}	5.625/30.33 = 0.1855

n-C_4H_{10}	5.625/30.33 = 0.1855
i-C_4H_{10}	6.19/30.33 = 0.2041
n-C_5H_{12}	0.96/30.33 = 0.0317

Composition of Liquid

Feed Composition	K @ 81°F	Mol Fraction x = y/K
C_2H_6	6.6	0.3651/6.6 = 0.0553
C_3H_8	1.78	0.2137/1.78 = 0.1201
n-C_4H_{10}	0.54	0.1855/0.54 = 0.3435
i-C_4H_{10}	0.77	0.2041/0.77 = 0.2651
n-C_5H_{12}	0.16	0.0317/0.16 = 0.1981

This should be = 1.00. Inaccuracy in reading K values probably accounts for most of the difference.

10.18. Determination of Distillation Operating Pressures

Determining the proper operating pressure for a distillation system, whether using trays or packed column, requires examining the conditions by following the pattern of Figure 10-33 [149]. It is essential to establish condensing conditions of the distillation overhead vapors, and any limitations on bottoms temperature at the estimated pressure drop through the system. Preliminary calculations for the number of trays or amount of packing must be performed to develop a reasonable system pressure drop. With this accomplished, the top and bottom column conditions can be established, and more detailed calculations performed. For trays this can be 0.1 psi/actual

tray to be installed [149] whether operating at atmospheric or above, or 0.05 psi/tray equivalent for low vacuum (not low absolute pressure).

Figure 10-33

Algorithm for establishing distillation column pressure and condenser type.

Source: Henley, E. J. and Seader J. D., Separation Process Principles, 2nd Edition, John Wiley © p. 265 (2006)].

Because low-pressure operations require larger diameter columns, selected pressures should only be as low as required to accomplish the separation.

For high vacuum distillation, Eckles *et al.* [150] suggest the use of a thin film, or conventional batch process, for industrial installations. However, many tray and packed columns operate at pressures as low as 4 mm Hg. abs. Eckles [150] suggests that “high vacuum” be taken as 5 mm Hg; molecular distillation as 0.3−0.003 mm Hg pressure, and unobstructed path distillation as 0.5−0.02 mm Hg. These latter two can be classed as evaporation processes. Eckles' [150] rules of thumb can be summarized:

1. Do not use a lower pressure than necessary, because separation efficiency and throughput decrease as pressure decreases.
2. The requirement to select a bottoms temperature to avoid overheating heat sensitive materials may become controlling.
3. When separating volatile components such as a single stream from low-volatility bottoms, use a molecular or unobstructed path process of either thin film or batch type.

4. When separating a volatile product from volatile impurities, batch distillation is usually preferable.
5. Do not add a packed column to a thin film evaporator system, because complications may occur.

Note that good vapor-liquid equilibrium data for low-pressure conditions are very scarce and difficult to locate. However, they are essential for proper calculations. This is dealt with in References 151 and 152.

Studies with high-pressure distillation by Brierley [239] provide insight into some FRI studies, the effects of pressure on performance, and the impacts of errors in physical properties, relative volatility, *etc.* This work makes an important contribution to the understanding and setting of operating pressures.

10.19. Condenser Types from a Distillation Column

10.19.1. Total Condenser

In a total condenser all the overhead vapor is condensed to liquid. The heat load on the condenser is equal to the sum of the sensible heat above the dew point, plus the latent heat of vapor, plus the sensible heat in cooling liquid down to the bubble point. Condensation of liquid will commence at the dew point and end at the bubble point. (Any further removal would proceed due to cooling.) The condenser and accumulator pressure will be the total vapor pressure of the condensate. If an inert gas is present, the system total pressure will be affected accordingly. When using a total condenser, the condensed stream is subsequently split into two streams. One is returned into the column as reflux and the second leaves as distillate product.

10.19.2. Partial Condenser

The effect of a partial condenser is indicated in Figure 10-34 and is represented by the relations given in the rectifying and stripping

sections. The product is a vapor that is in equilibrium with the reflux to the column top tray, and hence the partial condenser actually serves as an “external” tray for the system; it should be treated as the top tray when applying the equations for total reflux conditions. This requires care in stepwise calculations of the column performance.

Figure 10-34

Total and partial condenser arrangements.

In a partial condenser, there are two general conditions of operation:

1. All condensed liquid is returned to the column as reflux, and all vapor is withdrawn from the accumulator as product. In this case the vapor $y_c = x_D$; Figure 10-5 and Figure 10-35.

Figure 10-35

Diagram of partial condenser; only a vapor product is withdrawn.

2. Both liquid and vapor products are withdrawn, with the liquid reflux composition being equal to liquid product composition. On an equilibrium diagram the partial condenser liquid and vapor stream's respective compositions are in equilibrium, but only when combined do they represent the intersection of the

operating line with the 45° slope (Figure 10-35).

Example 10-13: Calculation of Partial Condensation of a multicomponent feed

A multicomponent vapor feed (F) is partially condensed to produce a vapor (V) and a liquid (L). Mole fractions of components in the vapor feed are designated z, those in the vapor y, and those in the liquid x. The vapor-liquid equilibrium for any component is given by y = Kx. Assuming that the vapor and the liquid leave the partial condenser in equilibrium, show that for any component:

It has been calculated that at 18.9 bar and 55°C, one third of the vapor feed will be condensed.

Check this calculation for the feed composition and K values at 18.9 bar and 55°C given as:

Component		
z	0.0713	0.8978
K	2.65	1.0

Solution

Mass balance at the top of the vessel:

(10-156)

Component balance:

(10-157)

Since the vapor phase is in equilibrium with the liquid phase at the top of the vessel:

(10-158)

(10-159)

Substitution of Eq. (10-159) into Eq. (10-157) gives:

(10-160)

or

(10-161)

(10-162)

Let F = 3 kmol/h

Since 1/3 of the vapor feed is condensed, then

L = 1/3 of 3 kmol/h = 1 kmol/h

V = F – L = 2 kmol/h

To check the total mole fraction of the components at the top of the column, determine the mole fractions of individual components from Eq. 10-162, *i.e.*

Composition	**Feed Mol. Frac., z**	**Fz**	**Fz/V**	**K @ 55°F**	
C_2H_6	0.07135	0.2139	0.10695	2.65	0.1887
n-C_3H_8	0.8978	2.6934	1.3467	1.00	0.5000
n-C_4H_{10}	0.0309	0.0927	0.04635	0.41	1.2195

The condensing temperature range in a total condenser is dependent on the dew and bubble points, cooling water limitations, fouling and the pressure of the distillation column. The reflux temperature is also affected by the pressure control scheme and the presence of excess capacity for cooling. The control aspect example can be referred to when submerged tubes result from some pressure control schemes. The degree of sub cooling can be as much as 50 to 200°F.

10.20. Effect of Thermal Condition of Feed

The condition of the feed as it enters the column has an effect on the number of trays, reflux requirements and energy requirements for a given separation. Figure 10-36 illustrates the possible situations, i.e., subcooled liquid feed, feed at the boiling point of the column feed tray, part vapor and part liquid, all vapor, or superheated vapor. The

thermal condition of the feed stream is designated as “q”; this is approximately the amount of heat required to vaporize one mol of feed at the feed tray conditions, divided by the latent heat of vaporization of the feed.

Figure 10-36

Operating characteristics of distillation columns.

For a single feed stream column with a total condenser and a partial reboiler as shown in Figure 10-5, a mass balance on the rectifying section of the more volatile component is:

(10-163)

A mass balance on the stripping section is:

(10-164)

Assuming constant molar overflow, at the feed plate the point at which the top operating line intersects the bottom operating line is:

(10-165)

Subtracting Eq. 10-163 from Eq. 10-164 gives:

(10-166)

From the overall mass balance around the entire column (see Figure 10-5), where:

(10-167)

(10-168)

Equation 10-168 gives a form of the feed equation since L, , V, , F and x_F are constant. It represents a straight line (the feed line) on a McCabeThiele diagram. The case where the feed flashes into the column to form a vapor and a liquid phase is illustrated in Figure 10-37. Part V_F vaporizes while the remainder is liquid, L_F. From Equation 10-168, is the change in liquid flow rates at the feed stage. That is:

(10-169)

Figure 10-37

Twophase feed.

The change in vapor flow rate is:

(10-170)

Equation (10-168) becomes:

(10-171)

Equation (10-171) represents the flashing of the feed into the column, and can be written in terms of the fraction vaporized as:

(10-172)

or in terms of the fraction of remaining liquid,

(10-173)

The slope of a line from the intersection point of the feed composition, x_F, with the 45° line on Figure 10-5 is given by $q/(q - 1) = -q/(l - q)$. Physically this gives a good approximation of the number

of mols of saturated liquid that will form on the feed plate by the introduction of the feed, although under some thermal conditions the feed may vaporize liquid on the feed plate rather than condense any.

Liebert [218] studied feed preheat conditions and the effects on the energy requirements of a column. This is an essential consideration in the efficient design of a distillation system. As an alternative to locating the "q" line, any value of x_i may be substituted in the "q" line equation below, and a corresponding value of y_i determined, which when plotted will allow the "q" line to be drawn in. This is the line for SV − I, V − I, PV − I, BP − I and CL − I of Figure 10-36.

(10-174)

Equation 10-173 or 10-174 is termed the q-line equation. It is located graphically as follows. When $x_i = x_F$, Equation 10-174 gives $y = x_F$. Thus, the q-line passes through the point (x_F, x_F). The slope of the q-line is then calculated on one other convenient point.

A more general definition of q is applied to determine the values of q for subcooled liquid and superheated vapor:

> q = enthalpy change to bring the feed to dewpoint vapor temperature divided by enthalpy of vaporization of the feed (dewpoint minus bubble-point), that is,
>
> (10-175)
>
> where
>
> H_F = feed enthalpy at saturated vapor temperature, kcal/kg, Btu/lb mol.

h_F = feed enthalpy (liquid, vapor or twophase), kcal/kg, Btu/lb mol.
h = liquid enthalpy, kcal/kg, Btu/lb mol

For a subcooled liquid feed, Equation 10-156 becomeswhere

(10-177)

and

(10-178)

or

(10-179)

For a superheated vapor, Equation (10-175) becomes:

(10-180)

where and are the liquid and vapor molar heat capacities respectively, is the molar enthalpy change from the bubble point to the dew point, and and are the feed, dewpoint

and bubble-point temperatures, respectively of the feed at the column operating pressure. As q changes from a value greater than 1 (subcooled liquid) to a value less than 0 (superheated vapor), the slope of the q-line, q/(q-1), changes from a positive value to a negative value and then back to a positive value. Figure 10-38 shows the effect of thermal condition of the feed on slope of the q-line. For a saturated liquid feed, the q-line is vertical and for saturated vapor, the q-line is horizontal as shown in Figure 10-38.

Figure 10-38

Location of q-lines.

The relationship between the various types of feed and the slopes of the feed condition is shown in Table 10-22.

Table 10-22 Feed conditions

Type of feed	**T**	**h_F**
Subcooled liquid		
Saturated liquid		h
Liquid-vapor		

vapor mixture		
Saturated vapor		H_F
Superheated vapor		
T_{BP} = bubble point of feed; T_{DP} = dew point of feed.		

10.21. Effect of Total Reflux, Minimum Number of Plates in a Distillation Column

Total reflux exists in a distillation column, whether a binary or multicomponent system, when all the overhead vapor from the top tray or stage is condensed and returned to it. A column may be brought to equilibrium at total reflux for test purposes or for a temporary plant condition, which requires temporary discontinuation of operation. Rather than shut down, drain, and then re-establish operating conditions later, it is usually more convenient to maintain a total reflux condition with no feed, no overhead and no bottoms products or withdrawals. This approach also requires less energy in the form of reboiler heating and condenser coolant

Conditions of total liquid reflux in a column coincide with the minimum number of plates required for a given separation. Under such conditions the column has zero yield of product, infinite heat requirements, and $L_s/V_s = 1.0$, as shown in Figure 10-36. This is the limiting condition for the number of trays, and is a convenient measure of the complexity or difficulty of separation.

10.21.1. Fenske Equation: ShortCut Prediction of Overall Minimum Total Trays In A Column With Total Condenser

(10-181)

This includes the bottoms reboiler being equivalent to a tray in the system.

Refer to tabulation below.

N_{min} includes only the required trays in the column itself, and not the reboiler.

$\alpha_{avg} = (\alpha_{lk/hk})$ avg.

D refers to overhead distillate; B refers to bottoms

(10-182)

This applies to any pair of components. Ludwig [271] suggested adding +1 theoretical tray for the reboiler, thus making the total number of theoretical trays a conservative estimate. Converting to the actual number of trays requires the calculated tray efficiency:

(10-183)

For a condition of overall total trays, allowance is to be made for feed tray effect; *add one more theoretical tray to the total*. As demonstrated below, allowance should be made for the reboiler and condenser.

			To Cond
			+1

		N_{min}	+1
S_m		N_{min}	+1
		N_{min}	+0

The approach recommended here is not in agreement with Van Winkle [74], who assumes that the reboiler and partial condenser are included in the overall calculation for N_{min}.

Various average values of α for use in these calculations are suggested in the following section on "Relative Volatility". Because the feed tray is essentially non-effective, an additional theoretical tray may be added to allow for this. This can be conveniently solved by the nomographs [21] of Figure 10-39 and Figure 10-40. The minimum number of trays in the rectifying section can, if necessary, be calculated by the Fenske equation, substituting the limits of x_{Fl} for x_{Bh} and x_{Bl}, and the stripping section can be calculated by difference.

Figure 10-39

Approximate solution for N and L/D in distillation of ideal binary mixtures.

Used by permission, Faasen, J.W., Industrial & Eng. Chemistry, V. 36 (1944), p. 248. The American Chemical Society, all rights reserved.

Figure 10-40

Minimum reflux at infinite theoretical plates.

Used by permission, The American Chemical Society, Smoker, E. H., Ind. Eng. Chem V. 34 (1942), p. 510, all rights reserved.

From Fenske's equation, the minimum number of equilibrium stages at total reflux is related to their bottoms (B) and distillate or overhead (D) compositions using the average relative volatility, see Equation 10-161.

To solve for the component split [100] in distillate or bottoms:

(10-184)

where

S_m = total number of calculated theoretical trays at total reflux, from Equation 10-181

$x_{lk} = x_{LK}$ = liquid mol fraction of light key

$x_{hk} = x_{HK}$ = liquid mol fraction of heavy key

lk − hk = LK − HK = average relative volatility of column (top to bottom)

Clearly, a column cannot operate at total reflux and produce net product. Hence, a reflux ratio of about *1.1 to 1.5 times the minimum reflux* will usually give practical results. As the reflux ratio is reduced, the number of theoretical trays increases and an economic balance is therefore involved.

10.22. Relative Volatility (α) Separating Factor in a Vapor-Liquid System

The relative volatility is a measure of the ease of separation, and is the volatility separation factor in a vapor-liquid system, i.e., the volatility of one component divided by the volatility of the other and can be defined as:

(10-185)

It represents the ratio between the tendency to vaporize of the two components. It also quantifies the tendency for one component in a liquid mixture to separate from the other upon distillation. The term can also be expressed as the ratio of vapor pressure of the more volatile to the less volatile component in the liquid mixture, and therefore $\alpha_{1,2}$ is always ≥1.0. If $\alpha_{1,2}$ is high, component 1 has a greater tendency to vaporize (i.e. is more volatile than component 2) and it will be easy to achieve separation by vaporizing one from the other distillation. However, when component 2 has as high a tendency to vaporize as component 1, the relative volatility will approach unity; the components will then be difficult to separate by distillation. If under the specific conditions , the components cannot be separated by distillation. The greater the value of α the easier will be the desired separation.

Relative volatility can be calculated between any two components in a mixture; binary or multicomponent. One of the substances is chosen as the reference to which the other component is compared.

Definition of Relative Volatility: Relative Volatility of Component 1 with respect to component 2:

(10-186)

where

1, 2, *etc*. are component identification
p = vapor pressure of a component
x = liquid mol fraction of a component
y = vapor mol fraction of a component
π = system total pressure, absolute

Partial pressure:

(10-36)

When temperature is constant and at equilibrium for a homogeneous mixture (such as an azeotrope), the composition of the liquid is identical with the composition of the vapor, thus $x_i = y_i$, and the relative volatility is equal to 1.0.

(10-38)

(10-39)

where

i = compound identification
r = reference compound

As previously discussed, the charts of K values that are available, apply primarily to hydrocarbon systems. Reference 79 presents other important data on K value relationships. Refer to Figure 10-11a and Figure 10-11b for charts with pressure effects included. (These are not ideal, but practical charts.)

(10-39)

For multicomponent mixtures [79, 59]:

(10-187)

where

1, 2, 3, 4, … are components in a multicomponent mixture.

$\alpha_{1/2}$ = relative volatility of component 1 with respect to component 2.

$\alpha_{3/2}$ = relative volatility of component 3 with respect to component 2.

(10-188)

(10-189)

(10-190)

(10-191)

For a binary system with constant relative volatilities:

(10-192)

or

(10-193)

Winn [99] proposes a modification to allow for temperature variation effects on relative volatility. The method does not apply to mixtures forming azeotropes or at conditions near the critical point. Kister [94] proposes:

(10-194)

α can vary with temperature, so an average α value should be used between top and bottom temperature.

When b_{lk} and $\beta_{lk/hk}$ are constants at a fixed or constant pressure, but evaluated for the light (1) and heavy (h) keys at top and bottom

temperatures, their relationship is [94]:

(10-195)

Winn's equation reduced to Fenske's at $b_{lk} = 1.0$ and

(10-196)

Example 10-14: Determination of Minimum Number of Trays by Winn's Method (used by permission [99])

The minimum number of trays necessary to debutanize the effluent from an alkylation reactor will be calculated. The feed, products, and vapor-liquid equilibrium constants of the key components at conditions of temperature and pressure corresponding to the top tray and reboiler are shown in Table 10-23.

Table 10-23 Data for Alkylation of Deisobutanizer; Example 10-14, using Winn's Method.

				Equilibrium K's	
Component	**Feed, moles**	**Overhead, moles**	**Bottoms, moles**	**Top tray**	**Reboiler**
Ethylene	1	1			
Ethane	2	2			
Propane	48	48			
Isobutene	863	848	15	0.94	3.55
n-Butane	132	71	61	0.70	3.00
Isopentane	33		33		
n-Pentane	5		5		

Alkylate	277		277		
	1361	970	391		
Used by permission, Winn. F.W., *Pet. Ref.* V. 37, No. 5 (1958). P.216. Gulf Pub. Co.					

The constants β and b are evaluated using Equation 10-195 as follows:

$$0.94 = \beta\,(0.70)^b$$
$$3.55 = \beta\,(3.00)^b$$

Divide to solve for value of b. Then:

$$3.78 = (4.29)^b$$
$$b = 0.913$$
$$\beta = 1.301$$

By use of Winn's Method [99] for product rates:

(10-197)

(10-198)

where

B = mols of bottoms
b = exponent in Equation 10-195
D = total mols of overhead product
n = minimum number of equilibrium trays in tower
K = y/x = vapor-liquid equilibrium ratio for a component
L = mols of a component in liquid phase
P = vapor pressure, psia
T = absolute temperature, °R

V = mols of a component in vapor phase
W = total mols of bottoms product
x = mol fraction of a component in liquid phase
y = mol fraction of a component in vapor phase
α = relative volatility
β = constant in Equation 10-194
π = total pressure, psia
L = total mols in liquid phase
V = total mols in vapor phase

subscripts or superscripts:
D = distillate
B = bottoms
(′) = heavy key component
1, 2… = tray number

The minimum number of theoretical stages is calculated as follows:

This is exactly the number of stages obtained by tray-to-tray calculations with the K correlation of Winn [235]. The minimum number of stages by the Fenske equation, with a geometric average α of 1.261, is 16.8. Hence the Fenske equation gives an answer greater by 2.3 stages, *i.e.* 16%.

For ideal systems following Raoult's Law; relative volatility , ratio of partial pressures.

For a binary distillation, α is calculated at top and bottom conditions and a geometric mean used where the differences are relatively small.

(10-199)

Kister [94] recommends that the determination of α for calculation is:

(1) α_{avg} = α evaluated at $T_{avg} = (T_{top} + T_{Bot})/2$

(10-200)

(2) α_{avg} = α at feed tray temperature

(3) Winn's [99] method previously discussed

For hydrocarbon systems; the following is often used [65]:

(10-201)

where

i = any component

r = component to which all the relative volatilities are referred

K_i = equilibrium distribution coefficient for component, i

K_r = equilibrium distribution coefficient for component to which relative volatilities are referred

For values of $\alpha \simeq 1.0$, care must be taken in establishing data, as a small change in the value of α_{avg} may double the number of trays.

The exact procedure is to estimate a temperature profile from top to bottom of the column and then calculate α for each theoretical tray or stage by assuming a temperature increment from tray to tray. For many systems this, or some variation, is recommended to perform accurate separation calculations.

(10-202)

For nonideal systems:

(10-203)

The vapor–liquid equilibrium relationship may be determined from:

(10-204)

By assuming values of x_l, the corresponding y_1 may be calculated.

For hydrocarbon systems, where $K_i = y_i/x_i$, then,

(10-205)

(10-206)

Example 10-15: Construction of Boiling Point Curve and Equilibrium Diagram for Benzene-Toluene Mixture

Using the vapor pressure data for benzene and toluene [59]:

1. Construct a boiling point diagram at a total pressure of 760 mm Hg, Figure 10-41.

Figure 10-41

Boiling point diagram for Example 10-15. Benzene-toluene mixture, total pressure = 760 mm Hg.

2. From the boiling point diagram construct the equilibrium x-y curve for a total pressure of 760 mm Hg, Figure 10-42.

Figure 10-42

VLE diagram for the mixture benzene and toluene at 760 mm Hg total pressure.

Vapor pressure data as read from tables or graphs;

Temperature°C	Vapor Pressure Benzene, mm Hg	Vapor Pressure Toluene, mm Hg

	mm Hg	mm Hg
80	760	280
90	1,000	410
100	1,320	550
110	1,740	740
111.5	1,760	760

Solution

Use Raoult's and Dalton's Laws:

$p_B = P_B x_B$

$p_T = P_T x_T$

Then:

$x_T = 1 - x_B$

$y_T = 1 - y_B$

$y_B = p_B/\pi$

where

P_B = vapor pressure, benzene

P_T = vapor pressure, toluene

π = Total pressure

Temp, °C	**(P_B – P_T)**	**(π – P_T)**	**x_B**	**x_T**		**y**
80	480	480	1.0	0	1.0	0
90	590	350	0.593	0.407	0.780	0.
100	770	210	0.273	0.727	0.474	0.
110	1,000	20	0.0	0.980	0.0457	0.
111.5	1,000	0	0.0	1.0	0.0	1.

Plot values of x_B, x_T, y_T, and y_B on Figure 10-42.

Example 10-16: Repeat Example 10-15 using K-values. Refer to basis of Example 10-14.

Using the data from Reference 59 (pp. 221, and 233):

Temp. °C	**K_T**	**Relative volatility $\alpha_{B/T}$***	**$K_B = \alpha K_T$**	**$1 - K_T$**	**$(K_B - K_T)$**
80	0.37	2.65	0.981	0.63	0.611
90	0.53	2.48	1.314	0.47	0.784
100	0.73	2.39	1.745	0.27	1.015
110	0.97	2.35	2.28	0.03	1.31
111.5	1.0	2.35	2.35	0.0	1.35
*Read from chart [59]					

$X_B = 1 - K_T/(K_B - K_T)$	**$x_T = 1 - x_B$**	**$y_B = K_B x_B$**	**$y_T = K_T x_T$**
* 1.031	0.0	* 0.981	0.0
0.60	0.40	0.789	0.212
0.266	0.734	0.464	0.535
0.0229	0.977	0.0523	0.946
0.0	1.0	0.0	1.0
*Note: If graphs could be read accurately, these values would equal 1.0.			

Procedure

1. Read K value for toluene from tables or charts.
2. Read α values for benzene/toluene from Reference 59.
3. Calculate K (benzene) from: $\alpha_{B/T} = K_B/K_T$
4. Calculate $x_{benzene}$:

   $y_B = K_B\, x_B$

   $y_T = K_T\, x_T$

$$\Sigma 1.0 = K_B\, x_B + K_T\, x_T = K_B\, x_B + K_T\,(1 - x_B)$$
$$= x_B\,(K_B - K_T) + K_T$$
$$x_B = (1 - K_T)/(K_B - K_T)$$

5. Calculate $y_{benzene}$: $y_B = K_B\, x_B$
6. Plot boiling point diagram, see Figure 10-43.

Figure 10-43

Boiling point diagram for benzene-toluene mixture using K values, total pressure 760 mm Hg for Example 10-16.

7. Plot x–y diagram, see Figure 10-44.

Figure 10-44

x-y diagram for benzene in benzene-toluene mixture, 760 mm Hg pressure, based on K-values, Example 10-16.

Example 10-17: Calculation of Flash Vaporization of a Hydrocarbon Liquid Mixture

What fraction of a liquid mixture containing 10 mole % propane, 65% n-butane and 25% n-pentane would be vaporized in a flash vaporization process at a temperature of 40°F and a pressure of 600

mm Hg abs?

The following vapor pressure data for 40°F are available:

Propane	3,800 mm Hg
n-Butane	820 mm Hg
n-Pentane	190 mm Hg

Solution

Assume Raoult's Law is applicable. At a total pressure, π, the temperature of flash must be between the dew point and bubble point.

Raoult's Law:

$p_1 = P_1\ x_1\ K_1 = y_1/x_I\ \pi\ y_1 = P_1\ x_1\ y_1/x_1 = P_1/\pi = K_1$			
	Mol%	**Vapor Press.mm Hg = P_1**	
C_3H_8	10	3,800	6.34
nC_4H_{10}	65	820	1.37
nC_5H_{12}	25	190	0.317
8.64			
37.6			
+ 6.0			
V = 52.24 NOT a check, reassume			

and recalculate. See Figure 10-45 for plot of results and the resultant extrapolation. This type chart serves as a guide to reduce the number of "guesses" to reach an acceptable solution. After several

assumptions:

1. Assume feed = 100 mols
2. Assume L = 30
3. Then: V = 100 − 30 = 70 mols

X = mols of component i in vapor plus mols of component i in liquid divided by the total mols of feed (liquid + vapor)

F = mols of feed

Figure 10-45

Extrapolation curve to determine approximate value of “V” for Example 10-17.

Following the same headings as in the previous table it continues:

	Mol%	**FX**	**Vapor Press.**	
C_3H_8	10	10	3,800	0.429
n-C_4H_{10}	65	65	820	0.429
n-C_5H_{12}	25	25	190	0.429
		*y ; mol. Frac.		
9.38		0.135		
49.4		0.711		
10.63		0.153		
69.41		0.999		

These values are sufficiently close for most calculations. Therefore,

after several trial-and error calculations the results indicate that after flashing, there would be approximately 70% vapor of the above composition and 30% (mol) liquid.

10.23. Rapid Estimation of Relative Volatility

Wagle [92] presents an estimation method for the average relative volatility of two components, related to their normal boiling points and latent heats of vaporization within the temperature range of their individual boiling points:

(10-207)

where

α = relative volatility between the two components in the temperature range T_{bl} to T_{b2}

T_{bl} = normal boiling point of Component l, K

T_{b2} = normal boiling point of Component 2, K

L_l = latent heat of vaporization for Component 1 at T_{bl}, kcal/kmole

L_2 = latent heat of vaporization for Component 2 at T_{b2}, kcal/kmole

If a compound's latent heat is not known, it can be estimated from the normal boiling points and molecular weight.

Example 10-18: Estimation of Relative Volatility by Wagle's Method [92] (used by permission)

The average relative volatility of benzene and toluene can be determined using the following data: T_{bb} = 353.3 K, T_{bt} = 383.8 K, L_b = 7,352 kcal/kmole, and L_t = 7,930 kcal/kmole (where the subscripts b and t denote benzene and toluene, respectively). Substituting these values into Equation 10-207 above gives:

This compares with a value of 2.421 for α determined using vapor-pressure/temperature charts.

10.24. Estimation of Relative Volatilities under 1.25 (α < 1.25) by Ryan [271]

For key components with relative volatilities <1.25, a McCabeThiele diagram is difficult to create, because the stage equilibrium data and the operating line data are very close to the x-y diagonal line, and the graph is indecipherable. Ryan [271] recently developed an alternative diagram in which the transformation spreads out the region close to the diagonal over the whole height of the plot. This approach provides more insight into the separation of isobutane and n-butane as shown in Table 10-24.

Table 10-24 Components with Relative Volatilities less than 1.25

Components	Relative volatility, α
Butane splitters in natural-gas–liquid plants and refineries	about 1.25
Deisobutanizers in butylenes alkylation units in refineries	about 1.25
Argon columns in cryogenic air plants	
Columns to separate ortho, para- and metaxylene in petrochemical plants	about 1.15
Propane-propylene splitters	about 1.1
Ethylbenzene-xylene superfractionators	about 1.05

The procedure for creating an alternative diagram is as follows:

1. Plot (y – x) vs x instead of y on the y-axis.
2. The operating lines begin on the x-axis at x_B and x_D. Both slopes are [(L/V)-1]. The rectifying line slope will be negative and the stripping line slope positive.
3. The q-line starts on the x-axis at x_F. The value of q is the same as for a conventional McCabeThiele calculation. The slope of the q-line in the graph is the McCabeThiele slope minus 1. Therefore for a feed at its bubble point, the q-line is vertical (slope = infinity) for the conventional McCabeThiele and Ryan calculations. For the dew point feed, the slope is 0 for the conventional McCabeThiele and −1 for Ryan.
4. The familiar steps for the determination of theoretical stages on the McCabeThiele diagram are modified on the Ryan plot. The vertical portions remain vertical, but the horizontal portions become slanted with a slope of −1. Figure 10-46 and Figure 10-47 show McCabeThiele and Ryan's plots respectively.

Figure 10-46

McCabeThiele diagram.

Figure 10-47

Revised McCabeThiele diagram.

10.25. Estimation of Minimum Reflux Ratio: Infinite Plates

As the reflux ratio is decreased from infinity for the total reflux condition, more theoretical steps or trays are required for a given separation. When the limiting condition of Figure 10-48 is reached; where the operating line touches the equilibrium line, the number of steps from the rectifying to stripping sections becomes infinite.

Figure 10-48

Fractionation of binary mixture at minimum reflux condition.

If the operating lines of Figure 10-48 intersect at x_c, y_c outside or above the equilibrium line when insufficient reflux is used, the separation is impossible.

This graphical representation is easier to use for nonideal systems than the calculation method. This is another limiting condition for column operation, i.e., below this ratio the specified separation cannot be achieved even with an infinite number of plates. This minimum reflux ratio can be determined graphically from Figure 10-48, as the line with smallest slope from x_D intersecting the equilibrium line at the same point as the “q” line for mixture following Raoult's Law.

External reflux ratio = L/D

Slope of line from x_D:

(10-208)

L/V = internal reflux ratio

For nonideal mixtures the minimum L/V may be as indicated in Figure 10-36 (abnormal equilibrium), and hence not fixed as indicated above.

Figure a 10-40 presents a convenient and acceptably accurate nomogram by Smoker [66].

(10-209)

where x_c and y_c are coordinates of intersection of the minimum reflux "operating" line with the equilibrium curve. At boiling point $x_c = x_F$.

Underwood's algebraic evaluation [73] for minimum reflux ratio is applicable to handling ideal, or near ideal, systems:

For a liquid at its bubble point, q = 1.0

(10-210)

For high-purity separations where , with reasonably well balanced columns (where

) and moderate values of ,

(10-210a)

For a vapor feed with no superheat, q = 0

(10-211)

For the general case, the relation is more complex in order to solve for $(L/D)_{min}$

(10-212)

Short *et al.* [230] discuss minimum reflux ratio estimation for complex fractionators.

10.26. Calculation of Number of Theoretical Trays at Actual Reflux

The Gilliland correlation [23], Figure 10-49a, has proven satisfactory for many binary as well as multicomponent mixtures over a wide range of reflux ratios and number of theoretical plates. Many systems appear to be economically designed for and with actual reflux ratios of 1.2 to 1.5 times at minimum. This is inapplicable to systems of greatly varying relative volatility. A Ponchon-Savarit or enthalpy method must then be followed.

Figure 10-49a Correlation of theoretical plates with reflux ratio.

Eduljee [84] suggests an easier equation to replace the Gilliland plot. The data input is the same. For tray towers:

(10-213)

(10-214)

(10-215)

where

S_t = theoretical actual trays at actual reflux ratio, L/D, including overhead total condenser and reboiler

Y_T = correlation expression similar to Gilliland's

X = correlation expression similar to Gilliland's

R = reflux ratio, L/D, where L is liquid returned to the column in mols/hr

D = distillate rate in mols/hr

L = liquid returned to column, mols/hr

subscripts:

h = heavy

min = minimum
P = for packed towers
T = for tray column

After calculating X, and solving for Y using Equation 10-213, then solve for the theoretical trays, S_t, at the actual selected reflux ratio (L/D) from the equation for Y. The equation appears to represent several reliable data references. A review of alternative equations to replace the Gilliland plot is provided later in this book.

For packed towers, the corresponding relation for theoretical trays is [84]:

(10-216)

and,

(10-217)

where

NTU = total number of transfer units

, where α_h taken as 1.0

Mapstone [122] and Zankers [123] developed the chart shown in Figure 10-49b to follow Figure 10-49a to allow for a rapid evaluation of Gilliland's equation for the number of theoretical plates at any reflux, minimum number of theoretical plates and minimum reflux ratio. The accuracy appears generally satisfactory for initial industrial design. It can therefore be a useful time-saving device when

evaluating a range of values. For another interesting attempt to improve the Gilliland plot by use of equations, refer to Reference 136.

Figure 10-49b Chart for reflux vs. trays. Use this nomogram for Gilliland's calculations for number of theoretical plates/trays.

Used by permission, Mapstone, G. E., Hydrocarbon Processing, V/47, No. 5 (1968), p. 169, Gulf Publishing Co., all rights reserved.

Example 10-19: Using Figure 10-49B to Solve The Gilliland Equation to Determine the Minimum Number of Theoretical Plates for Selecting An Actual Reflux Ratio (used by permission [122])

If the minimum reflux ratio is 2.0 and the minimum number of theoretical plates is 20, how many theoretical plates will be required if a reflux ratio 1.5 times the minimum is used?

Solution

The required reflux ratio, R = 1.5 × 2.0 = 3.0

1. Connect 2.0 on left hand R_{min} scale with 3.0 on left diagonal R scale and extend to cut Tie Line 1.
2. Transfer this value across the central maze to Tie Line 2.
3. Connect this point on Tie Line 2 with 20 on the right hand S_{min} scale to cut the right diagonal S scale at 35 (calc. 34.9).

The number of theoretical plates required will be 35. If any three of the four variables, S, S_{min}, R, and R_{min} are known, this chart can be used by an analogous procedure to give the fourth.

where

S = theoretical plates at any reflux ratio

S_{min} = minimum number of theoretical plates

R = any reflux ratio
R_{min} = minimum reflux ratio

10.27. Identification of "Pinch Conditions" on a x-y Diagram at High Pressure

A column is referred to be "pinching" when the component balance line is too close to the equilibrium curve. This physically represents a situation where several stages in the column are accomplishing very little separation and are thus redundant. This phenomenon not only wastes stages, but could result in an oversized column. Additionally, inaccuracies in relative volatilities and enthalpies may bring the component balance line and the equilibrium curve closer or even cross the equilibrium curve earlier. This could increase the stage requirement, and if the additional stages required are unavailable, the column will not achieve its desired separation.

Wichterle *et al.* [91] identify that near the critical pressure point of the more volatile component, all systems exhibit a "pinch" phenomenon at high pressure, as shown in Figure 10-50 [91]. To obtain separation performance, the K-value data used must be accurate in this narrow range of separation. For hydrocarbon systems, as well as systems involving hydrogen, nitrogen, and methane, the data must be accurate and use of a general equation for the particular compound is inadequate. This is crucial for high accuracy in computer performance analysis.

Figure 10-50

Example of typical "pinch point" for critical region for high-pressure

distillation.

Used by permission, Wichterle, I., Kobayashi, R., and Chappelear, P. S., Hydrocarbon Processing, Nov. (1971) p. 233, Gulf Publishing Co., all rights reserved.

In an existing column, where the feed location is already in position, a pinch can be remedied by increasing the reflux and reboil. This invariably draws the component balance line and the equilibrium curve further apart, enlarging the steps and thereby permitting the required separation to be achieved with the existing feed location. The drawback is lower feed capacity, higher energy consumption, greater vapor and liquid flows through the column, and eventual higher operating costs. Therefore, it is essential to design a column away from the pinched region.

McCormick [97] presents a correlation for Gilliland's chart relating reflux ratio, minimum reflux ratio, number of theoretical stages, and minimum theoretical stages for a multicomponent distillation. Selection of a multiplier for actual reflux ratio over minimum reflux ratio is important. Depending on the complexity and analysis of the component's separation by the stages, the actual reflux ratio may vary from 1.2 to 1.5 to 2.0, dependent on economics. The proposed equation agrees satisfactorily with other methods, especially in the extreme ranges of Gilliland's plots [97], as well as the most used region.

(10-218)

(10-219)

Representative values of X calculated from Equation 10-219 are given in the following Table 10-25 for values of R_{min} and multiplier A. Actual reflux ratio values can be assumed, and the system tested for R_{min}.

Table 10-25 Operational Values of X Calculated via Equation 10-219 for a Range of Reflux Ratios*.

	R_{min}					**Multiplier A**					
	0.02	**1.05**	**1.07**	**1.10**	**1.12**	**1.15**	**1.17**	**1.20**	**1.30**	**1.40**	**1.50**
1	0.010	0.024	0.034	0.048	0.063	0.070	0.092	0.091	0.130	0.167	0.200
3	0.015	0.036	0.050	0.070	0.092	0.101	0.132	0.130	0.184	0.231	0.273
5	0.016	0.040	0.055	0.077	0.102	0.111	0.145	0.143	0.200	0.250	0.294
10	0.018	0.043	0.060	0.083	0.110	0.120	0.157	0.154	0.214	0.267	0.313
Total reflux	0.020	0.048	0.065	0.091	0.120	0.130	0.170	0.167	0.231	0.286	0.393

*Used by permission, McCormick, J. E., Chemical Engineering V. 95, No. 13 (1988), all rights reserved.

For actual versus minimum number of stages in a column:

(10-220)

(10-221)

(10-222)

where

A = parameter in correlating equation or multiplier on R_{min}

B = parameter in correlating equation

ln = natural logarithm

log = logarithm to the base 10

N = actual number of theoretical stages required for a given separation

N_{min} = minimum number of theoretical stages required for a given separation

R = external reflux ratio for a given separation

R_{min} = minimum external ratio for a given separation

$X = (R - R_{min})/(R + 1)$

$Y = (N - N_{min})/(N + 1)$

The following is a short approximation method for minimum reflux ratios for multicomponent mixtures [98]:

(10-223)

where

$(x_{FLK})_{eff} = x_{FLK}/(x_{FLK} + x_{FHK})$

n = number of components

R_{min} = minimum reflux ratio

x = liquid mol fraction

α_i = relative volatility of component i based on heavy key

α_{LK} = relative volatility of component, i, based on light key

subscripts:

avg = average

eff = effective
F = feed
FHK = heavy key in feed
FLK = light key in feed
i = component
LK = light key
HK = heavy key

Kister [94, 95] examines binary distillation systems with multiple feeds, one or more side products, one or more points of heat removal or addition, and various combinations.

10.28. Distillation Column Design

In designing a distillation column, the chemical/process engineer often is confronted with three basic tasks:

- Choice between tray or packed column
- Design of the column, and its ancillaries, *eg*. feed preheater, reflux condenser, reflux splitter, reboiler, *etc*.
- Rating problem

In the design of a plate column, it is a basic requirement to calculate how many plates/trays are required to achieve the desired degree of separation of the components from the specified feed conditions. This aspect relates to size and configuration of the equipment. In the rating, the performance of a column under various conditions with a fixed number of plates/trays after it is specified should be predicted.

10.28.1. Design Method for a Plate Column

The following is a design procedure using graphical solution.

Given:

Feed conditions

(i) Feed flow rate, F
(ii) Feed composition, z
(iii) Feed temperature or enthalpy or thermal condition, q
(iv) Column pressure
(v) Reflux temperature or enthalpy (usually saturated liquid)

Product Conditions

all with respect to light component

Reflux Ratio, R: Relative volatility, α, The vapor liquid equilibrium curve.

Procedure

1. Plot the vapor-liquid equilibrium curve from the data available at the column pressure. In terms of relative volatility:

(10-48)

where α is the geometric average relative volatility of the lighter (more volatile) component with respect to the heavier component (less volatile).
2. Draw the 45° – line, that is, x = y line.
3. Make a material balance over the column to determine the top and bottom compositions, x_B, x_D from the data given.
4. Locate the points x_B, x_D and z on the 45°-line. That is, draw perpendiculars on x-axis at points x_B, x_D and z. Their points of intersection with the 45°-line are the required locations.
5. Calculate the distillate and bottoms flow rates.
Distillate flow rate:

(10-224)

Bottoms flow rate:

(10-225)

6. The point of intersection of the two operating lines depends on the phase condition of the feed. The line on which the intersection occurs is referred to as the q line. The q-line is found as follows:

(i) Calculate the value of the ratio q given by

(10-226)

where the numerator in Equation 10-226 is the heat to make saturated vapor at the feed tray pressure from 1 mol of feed at its thermal condition, as fed.

(ii) Plot the q-line, slope = q/(q − 1), intersecting the diagonal at x_F (the feed composition).

7. Select the reflux ratio and determine the point where the top operating line extended cuts the y axis:

(10-227)

8. Draw in the top operating line, from x_D on the diagonal ϕ to the y-axis at the value ϕ.

9. Draw in the bottom operating line; from x_B on the diagonal to the point of intersection of the top operating line and the q-line.

10. Starting at x_D or x_B, step off the number of stages.

11. The design is complete.

Count the number of trays in the rectifying section.
Count the number of trays in the stripping section.
Sum the number of trays in both rectifying and stripping sections.

Note: The feed point should be located on the stage closest to the intersection of the operating lines.

The reboiler and a partial condenser (if used) act as equilibrium stages. Figure 10-5 shows the McCabeThiele diagram, and the method can be used for the design of columns with side streams and multiple feeds. The liquid and vapor flows in the sections between the feed and the take-off points are calculated and operating lines drawn for each section.

10.28.2. Continuous Fractionating Column

Rating Method

On completion of the column design, it should be checked to see whether it will perform under different conditions. This rating is an operating problem. Typically, the heat input rate to achieve a specified separation when the feed condition varies will be determined. Another type of rating problem can involve a trial-error iterative solution as follows:

(i) An assumption is made of one or more unknown variables.
(ii) On the basis of these assumed values, the design procedure is carried out.
(iii) The results are compared with the column specifications.
(iv) If they agree, no further assumption is required, and the estimated values of the unknown variables are acceptable; otherwise, another set of values are assumed and the design procedure repeated.

An essential difference between the design and rating methods is that the feed is not introduced on the optimum feed tray, but is obtained through design calculation. In general, the actual feed tray is not the optimum feed tray. Iterative procedures are essential to apply rating methods to the design of a new column. An initial estimate of the number of stages can be made using shortcut methods and programs used to calculate the product compositions. Calculations are repeated with revised estimates until a satisfactory design is achieved.

10.29. Simulation of a Fractionating Column

In a simulation exercise, the objective is to predict how much separation can be achieved with a given feed in a column, which has been commissioned or is in use. The number of stages and feed stage location are already fixed, as are the column diameter and the reboiler size; these fix the maximum vapor flow rate. There are various ways of specifying the remaining parameters. The desired composition of the more volatile component in the distillate and bottoms can be set; the requirement is to determine the external reflux ratio that will produce the separation. A check may also be made that the maximum vapor flow rate will not be exceeded. Alternatively, the distillate or bottoms compositions and reflux ratio are specified, and then the unknown compositions and the vapor flow rate are determine. Another alternative is to specify the heat load, in the reboiler, and the distillate or bottoms compositions thus requiring the reflux ratio and unknown product compositions to be determined and the vapor flow rate checked. In general, all variables that can be controlled are specified, and these are illustrated by Wankat [273].

Example 10-20

An equimolar binary mixture with a relative volatility of 2 is fed to a distillation column, which produces a distillate containing 97.5% of the more volatile component, and a residue containing 10% of the more volatile component. The feed is liquid at its boiling point and

80% of the liquid entering the reboiler is vaporized and returned to the column as boil-up.

(i) Estimate the number of theoretical plates provided by this column.

(ii) Check your answer in part (i) by reference to the graph of the Gilliland correlation.

Solution

The overall mass balance on the column.

(10-228)

The component balance on the more volatile component (MVC) is:

(10-229)

Assuming that F = 100 mols/hr.

$x_F = 0.5$, $x_D = 0.975$ and $x_B = 0.1$

Substituting the values of the components in Eq. 10-229 gives:

(10-230)

Rearranging Eq. 10-228 and substituting into Eq. 10-230 gives:

(10-231)

(10-232)

D = 45.71 mols/hr.

B = 54.29 mols/hr.

The overall mass balance on UOL is:

(10-233)

Component balance is:

(10-234)

or

(10-235)

(10-236)

Substituting Eq. 10-236 into Eq. 10-233 gives:

(10-237)

Equation Eq. 10-235 (UOL), assuming constant molar overflow, becomes:

(10-238)

The intercept on the UOL, is

(10-239)

Since the feed is liquid at its boiling point (i.e. q = 1) and 80% of the liquid entering the reboiler (L′) is vaporized to give the vapor (V′) leaving the reboiler and returned to the column.

Therefore at the bottom of the column, if L′ = 1.0 and V′ = 0.8.

The mass balance at the bottom of the column with the reboiler is:

(10-240)

or

(10-241)

The component balance at the bottom is:

(10-242)

or

(10-243)

Slope is

Using the McCabeThiele diagram of Figure 10-48, draw the lower operating line to intercept the “q” line; then draw the upper operating line and extend to the intercept I.

(10-244)

R = 3.43

From McCabeThiele diagram:

N +1 = 14 stages, hence

N = 13 stages.

R_{min}: from the points (X_D, X_D), a straight line is drawn that touches the VLE line with the “q” line and extended to the y-axis at a point .

(10-245)

Applying the Underwood Equation to find the minimum reflux ratio:

(10-210)

The Fenske method for the minimum number of theoretical stages at total reflux is:

(10-246)

From the Gilliland correlation, (Figure 10-49a and Figure 10-49b), the abscissa axis is:

(10-247)

The corresponding point on the ordinate axis of the Gilliland correlation is 0.31.

Therefore the number of theoretical stages can be determined from the ordinate by:

N = 12.69 stages (13 stages)

Example 10-21: Graphical Design of a Fractionation Column for Binary Systems [59]

The benzene–toluene example of Robinson and Gilliland [59] has been elaborated on and expanded after the advanced distillation course of Holland [25], Figure 10-51.

Figure 10-51

Equilibrium curve; benzene-toluene for Example 10-21 (curve data only).

Used by permission, Robinson, C. S. R. and Gilliland, E. R., Elements of Fractional Distillation, 4th Ed., McGraw-Hill Book Co. (1950), all rights reserved.

It is desired to separate an equimolal mixture of benzene and toluene into a top product containing 95 mol % benzene and a bottom product containing 95 mol % toluene. The distillation is to be carried out at atmospheric pressure. Use a total condenser.

A. Calculate the minimum reflux ratio if the feed is liquid at its boiling point.

B. Calculate the theoretical plates required if a reflux ratio (L/D) of 1.5 times the minimum is employed.

Feed = 50 mols benzene + 50 mols toluene

Overhead = 95% benzene

Bottoms = 95% toluene

Solution

Basis: Feed 100 mols/hr.

Material balance with respect to benzene (more volatile component, MVC):

$F = D + B$

Component balance on the MVC.

$0.50\ (100) = (0.95)\ (D) + 0.05B$

$45 = 0.90D$

$D = 45/0.90 = 50$ mols/hr overhead product

D = overhead product, mols/hr.

B = bottoms, mols/hr.

A. For a Feed at its Boiling Point:

Minimum L/V = 0.55 mol reflux/mol vapor up

Substituting:

$0.55\ L/D + 0.55 = L/D$

$0.45\ L/D = 0.55$

Reflux/Product = $L/D = 0.55/0.45 = 1.22$

The value of L/D minimum should be equal to:

The slight difference is probably due to inaccuracy in reading y_c = 0.70 from the equilibrium curve.

B. Number of Theoretical Plates at L/D = 1.5 times minimum:

Operating Reflux Ratio = (1.5) (1.25) = 1.878 = L/D

Slope of operating line at this reflux ratio:

From Graph, L/V = 0.653 was plotted based on feed at its boiling point, number of theoretical plates (stepwise graph) = 11.3

Now, to calculate the number of theoretical plates:

Rectifying section:

At: L/D = 1.878, D = 50 mols/hr. overhead

L = (1.878) (50) = 93.9 mols/hr. reflux to column

V = L + D

= 93.9 + 50 = 143.9 mols/hr. to vapor overhead

For a total condenser: $y_{top} = x_D = x_R = 0.95$

From the equilibrium curve at $y_t = 0.95$

then:

$x_t = 0.88$

$y_{(t-l)} = 0.651\ (x_t) + 0.331$

$y_{(t-l)} = 0.651(0.88) + 0.331 = 0.903$

$y_{t-1} = 0.903$, then x_{t-1} from equilibrium curve = 0.788

Now calculate y_{t-2}

$y_{t-2} = 0.651\ (0.788) + 0.331 = 0.844$

At $y_{t-2} = 0.844$, curve reads: $x_{t-2} = 0.69$

Then: $y_{t-2} = 0.651\ (0.69) + 0.331 = 0.780$

At y_{t-3}, curve reads: $x_{t-3} = 0.60$

Then: $y_{t-4} = 0.651\ (0.60) + 0.331 = 0.722$

At y_{t-4}, curve reads: $x_{t-4} = 0.52$ (Feed Tray)

Then: $y_{t-5} = 0.651\ (0.52) + 0.331 = 0.669$ (too far below feed)

Now go to stripping section curve:

The feed was at its boiling point:

$V_n = V_m = 143.9$

$B = F - D$

$= 100 – 50$

B = Bottoms = 50 mols/hr.

$L_m = B + V = 50 + 143.9 = 193.9$

Starting at t − 4 = feed tray:

$x_{t-4} = 0.52$

$y_{(feed-1)} = 1.35\ x_f - 0.0176\ (f-1)$

$y_{(f-1)} = 1.35\ (0.52) - 0.01736 + 0.685$

At $y_{t-1} = 0.685$, $x_{f-1} = 0.475$

Note: This is not too accurate due to switched operating line

equations before the feed compositions were reached, yet, one more calculation on the stripping line would have placed us below the feed plate composition. Hence a change in reflux ratio is necessary in order to split right at the feed composition [271].
Continuing:

$y_{f-2} = 1.35\ (0.475) - 0.01736 = 0.624$

From curve at $y_{f-2} = 0.624$

$x_{f-2} = 0.405$

$y_{f-3} = 1.35\ (0.405) - 0.01736 = 0.531$

From curve, $x_{f-3} = 0.32$

$y_{f-4} = 1.35\ (0.32) - 0.01736 = 0.416$

$x_{f-4} = 0.23$

$y_{f-5} = 1.35\ (0.23) - 0.01736 = 0.294$

$x_{f-5} = 0.15$

$y_{f-6} = 1.35\ (0.15) - 0.01736 = 0.186$

$x_{f-6} = 0.092$

$y_{f-7} = 1.35\ (0.092) - 0.01736 = 0.107$

$x_{f-7} = 0.05$ (The desired bottoms composition)

Total number of theoretical trays:

rectifying section = 4
feed tray = 1
stripping section = 7
total = 12 Trays

Example: 10-22 Variation of Theoretical Plates in a Column with Thermal Condition of Feed

Using the same operating reflux ratio as in Example 10-21, calculate the number of theoretical plates required for feed of the following

thermal conditions: Use Figure 10-52.

(a) q = 1.5

(b) q = 0

(c) q = − 1.5

A. For q = 1.5

Slope of "q" line = −q/(1 − q)

Figure 10-52

Equilibrium curve; benzene–toluene for Example 10-22 (curve data only).

Used by permission, Robinson, C. S. R. and Gilliland, E. R., Elements of Fractional Distillation, 4th Ed., McGraw-Hill Book Co. (1950), all rights reserved.

Substituting, slope =

Referring to calculations of Example 10-21, for an equimolal mixture of benzene and toluene in feed:

overhead product, D = 50 mols/l00 mols feed

calculate

where $x_D = 0.95$

$y_c = 0.774$*

$x_c = 0.59$*

$(L/D)_{min} = (L_R/D)_{min} = 0.956$ min. reflux ratio, reflux/ product

*Read from graph at intersection of "q" line for 1.5 and

minimum reflux operating line.

Slope of Operating Line at Minimum Reflux:

(Graph reads 0.59 but this depends upon accuracy of plot.)

Calculating

Actual Operating Line

Operating reflux ratio = (1.5) (L/D) = 1.5 (0.956)
= 1.432 reflux/product

Slope of actual operating line:

Graphically, read 13 steps or theoretical plates from the top plate through bottom reboiler (assuming a total condenser).

rectifying section	= 5
feed plate	= 1
stripping section	= 7 (includes reboiler)
	13 Plates including reboiler

To calculate this stepwise:

Operating line of rectifying section:

$L/V = 0.59$

$L/D = 1.432$, $D = 50$ mols product

$L = (50)\ 1.432 = 71.6$ mols liquid reflux

$V_r = L_r + D = 71.6 + 50 = 121.6$ mols

then:

$y_{n+1} = 0.59\ x_n + 50\ (0.95)/121.6$

$y_{n+1} = 0.59\ x_n + 0.39$, operating line equation

At top: $y_{n+1} = x_D = 0.95$

So: From equilibrium curve at $y_{n+1} = 0.95$, read the liquid in equilibrium, which is x_n (or top plate in this case) $x_n = x_{top} = 0.88$.

Now substitute this value $x = 0.88$ into the equation and calculate the vapor coming up from the first plate below the top ($t - 1$). Thus, if x_n = top plate, y_{n+1} = vapor from plate below top. Now, read equilibrium curve at $y_{(t-1)}$ and get $x_{(n+1)}$ or x_{t-1} which is liquid on plate below top. Then using x_{t-1}, calculate y_{t-2} (second plate below top, etc.). Then, read equilibrium curve to get corresponding liquid x_{t-2}. Continue until feed plate composition is reached, then switch to equation of stripping section and continue similarly until the desired bottoms composition

is reached.

Operating line of stripping section:

Because the feed is a supercooled liquid, $L_s/V_s \neq L_r/V_r$. From definition of "q":

$L_s = L_r + qF$

$L_s = 71.6 + (1.5)(100)$

$L_s = 221.6$

Also:

$121.6 - V_s = -50$

$V_s = 171.6$

Therefore,

Stripping section operating line:

$y_m = 1.29\, x_{m+1} - 0.291\, x_B$

$x_B = 0.05$

$y_m = 1.29\, x_{m+1} - 0.01455$

Use this equation as described above proceeding down from the feed plate crossover from the rectifying equation to the stripping equation.

B. For q = 0

This represents feed as all vapor (not superheated). Slope of “q” line:

This represents no change in overflow from the feed plate, and the increase in vapor flow is equal to the mols of feed.

where:

min. reflux ratio, reflux/product

Slope of operating line at minimum reflux:

Slope from graph = 0.688

Operating reflux ratio = (1.5) (2.14) = 3.21, reflux/product, $(L/D)_{op}$

Slope of operating line = $(L/V)_{op}$ = 3.21/ (3.21 + 1) = 0.763

No. of theoretical plates from graph	= 11

No. plates rectifying section	= 5
Feed plate	= 1
Stripping section	= 5 (includes reboiler)
Total	= 11 (includes reboiler)

Rectifying Section Equation for Operating Line

$L/V = 0.763$

$L/D = 3.21$

L = (3.21) (50 mol product. D) = 160.6 mols (reflux liquid)

$V_r = L_r + D = 160.6 + 50$

V_r = 210.6 mols vapor up column

then:

$y_{n+1} = 0.763\ x_n + 0.225$

Liquid Down Stripping Section

$L_s = L_r + qF$

$L_s = 160.6 + (0)$ (100 mols feed)

$Ls = 160.6 = L_r$

Vapor Up Stripping Section

$210.6 - V_s = 100$

$V_s = 110.6$ mols

Stripping Section Equation for Operating Line

$y_m = 1.452\ x_{m+1} - 0.0226$

Use these equations as described for the (a) part of problem in solving for number of theoretical plates stepwise.

C. For $q = -1.5$

This represents feed as a superheated vapor, and there is a decrease in liquid overflow from feed plate.

where

$(L/D)_{min}$ = 4.84 reflux/product

Slope of operating line at minimum reflux:

Actual Operating Line

Operating reflux ratio = (1.5) (4.84) = 7.26 reflux/product

Slope of actual operating line = (L/D) =

Graphically we read 8.5 total plates through to the bottom reboiler

Rectifying section	= 5
Feed plate	= 1
Stripping section	= 2.5 (includes reboiler)
Total	= 8.5 (includes reboiler)

Equations for Stepwise Tray to Tray Calculations Rectifying Section operating Line

L_r/V_r = 0.879

L/D = 7.26

$L_r = (7.26)\ (50) = 363$ mols liquid reflux

$V_r = L_r + D = 363 + 50 = 413$

$y_{n+1} = 0.879x_n + 0.115$

Liquid Down Stripping Section

$L_s = L_r + qF$

$L_s = 363 + (-1.5)\ (100) = 213$ mols liquid

Vapor up Stripping Section

$-V_s = 250 - 413$

$V_s = 163$ mo1s vapor

Stripping Section Operating Line

Use these equations as described for Part (a) in solving for theoretical plates.

10.30. Determination of Number of Theoretical Plates in a Fractionating Column by the Smoker Equations at constant relative volatility (α = constant)

Smoker [274] developed an analytical equation to determine the number of stages when the relative volatility is constant. This method can be used in preference to the McCabeThiele [225] graphical method because the accuracy of the latter depends on the care given to the construction of the plots. The graphical construction for the number of stages also becomes difficult at very low concentrations. Further, if the relative volatility ≃ 1.0, the number of stages required will be very large, and the diagram will be difficult to construct.

The Equations

Smoker's equations consider Raoult's law for a binary mixture, where α is a constant, and represent the vapor–liquid equilibrium relationship. From Equation 10-48:

(10-48)

Any operating line that is either above or below the feed plate can be represented by the general equation of a straight line:

(10-248)

Eliminating y from Equation 10-48 and 10-248 gives a quadratic equation in x:

(10-249)

(10-249)

For any particular distillation problem, Equation 10-249 will have only one real root k, between 0 and 1. Representing this root k, and substituting in Equation 10-249 gives:

(10-250)

k is the value of the x-ordinate at the point where the extended operating lines intersect the vapor–liquid equilibrium curve.

Smoker shows that the number of stages required is given by:

(10-251)

where

(10-252)

N = number of stages required effecting the separation represented by the concentration change from

to

;

and

;

(10-253)

(10-253)

k = composition of the liquid where the operating line intersects the equilibrium line

m = slope of the operating line between and

α = average relative volatility, assumed constant over to =

10.30.1. Application of Smoker's Method to a Binary Distillation Column

Applying Smoker's equation to actual distillation conditions with a single feed and no side streams.

Rectifying Section

For a rectifying column, the number of plates required to enrich from the feed composition, x_F to the distillate (product) composition, x_D at a fixed reflux ratio, R is obtained from:

(10-254)

and

(10-255)

also

(10-256)

where k is the root between 0 and 1

(10-257)

(10-258)

(10-259)

(10-260)

Stripping Section

For the stripping section, the bottoms composition, x_B, the feed, x_F, the overhead composition, x_D, and the reflux ratio, R all fix the operating line. The values of m and b are:

(10-261)

and

(10-262)

(10-263)

(10-264)

If the feed stream is not introduced at its bubble point, x_F is replaced by the value of x at the intersection of operating lines given by:

(10-265)

For distillation at total reflux R = ∞, m = 1, b = 0, k = 0, c = 1 and Equation 10-251 becomes:

(10-266)

Equation 10-251 simplifies to the equation derived by Underwood [73]. Smoker's equation is not limited to the range of low concentrations, but can be used for stage calculation over the entire range. In such instances, the equation is written twice, one for each section of the column. The average relative volatilities for the rectifying section and for the stripping section are determined.

Example 10-23 Use of Smoker's Equation to Determine the Number of Theoretical Stages and The Optimum Feed Stage for Binary Distillation

A column is to be designed to separate 1000 moles/hr of a binary mixture of benzene and toluene. The feed

contains 40% benzene and 60% toluene. A distillate that is 99% benzene and a bottoms that is 1% benzene are desired at a reflux ratio of 3 to 1. For this mixture, the average value of relative volatility (α) is 2.50. Estimate the number of equilibrium stages at this reflux ratio and the optimum feed stage location.

Solution

Rectifying Section

(10-256)

Solving

Since , Use k = 0.21847

(10-251)

= 7.88 stages.

Stripping Section

Feed is taken at its bubble point.

(10-261)

(10-256)

Solving

$= 0.5487$ or -0.00333

Since $0 < k < 1$, use $k = 0.5487$

Total number of stages is therefore:

$N_s = 7.88 + 7.79 = 15.67$, say 16

A Fortran program PROG103.for has been developed, which uses a quadratic equation to determine the number of stages in the rectifying and stripping sections of a distillate column. The program uses Smoker's equations involving two components to calculate the required number of stages. PROG103 was developed using Absoft v.10 Fortran compiler in the Microsoft run time Windows Environment (MRWE) application. The user need only create the data source file (e.g. DATA103.DAT) and run the executable PROG103.exe to generate the results. Both the source and executable codes can be downloaded from the Elsevier companion website. Table 10-26 shows typical input data and results. The required number of stages between benzene and toluene mixture is 16 and the optimum feed location is 8. Constructing the McCabeThiele diagram using Figure 10-42 a for the more volatile component (i.e. benzene) with the feed at its bubble point (q = 1), gives the number of stages as 18.

Table 10-26 Data Input and Computer Results for Calculating Equilibrium Number of Stages Using Smoker's Equation

Data name:	Data103.Dat	
2.5	2.5	
0.4	0.99	0.01
3		
Equilibrium number of stages by smoker's equation		
The relative volatility of component 1. :	2.500	
The relative volatility of component 2. :	2.500	
The mole fraction of the feed:	0.4000	

The mole fraction of the distillate:	0.9900
The mole fraction of the bottoms:	0.0100
The reflux ratio:	3.00
The number of stages in the rectifying section:	7.87
The number of stages in the stripping section:	7.76
Total number of stages:	15.63
Feed stage location:	7.8

10.31. The Jafarey, Douglas and McAvoy Equation: Design and Control [275, 276]

Operability and control are important considerations for any column. For example, it is common practice for operators to increase the reflux rate above its design value and thereby increase the purity of the overhead stream (assuming that the desired product arises from the overhead). Using this approach, the disturbances in the feed composition seldom cause the top product to fall below its quality specification. However, this advantage must be balanced by the excess steam supply needed to over-reflux the column. To solve problems of this type, and to develop better control systems for columns, Shinskey [277] developed an empirical correlation between the boilup ratio and the separation factor for a column, and Smoker [274] developed an exact analytical solution for binary distillation columns for the case of constant relative volatility.

Jafarey *et al.* [275, 276] derived an approximate equation for binary distillation by simplifying the solution of Smoker's equation. This equation can be used to predict the effect of disturbances on column performance, and is thus useful in computer and microprocessor control, where it can be applied to estimate the effect of disturbances and the control action required to compensate for them. The equation is:

(10-267)

where

N = Number of stages.

(10-267)

R = Reflux ratio.

q = Number of pound-moles of liquid formed on the feed stage when introducing 1 lb-mole of feed.

z = Mole fraction in the feed.

α = Relative volatility.

Example 10-24 Design and Control of a distillation column using The Jafarey, Douglas and McAvoy Algorithm

A benzene-toluene column normally operates as described in Example 10-23 with 30% vaporized. The algorithm manipulates boilup flow to control toluene purity. If the toluene purity is to be temporarily, increased from 90 to 95% and the benzene purity is to remain unaffected, to what the boilup flow rate to set? Assume that any boilup changes will be compensated for by reflux and distillate rate changes, such that the benzene purity remains unaffected. The relative volatility is 2.5.

Solution

The overall mass balance:

$$1000 = D + B$$

Component balance on the MVC (benzene)

$1000\ (0.4) = D \times 0.99 + B \times 0.1$

$400 = 0.99\ D + 0.1\ B$

$400 = 0.99\ (1000 - B) + 0.1B$

$B = 663$ lb mole/h

$D = 1000 - 663$

$= 337$ lb mole/h

In the rectifying section of the column

$V = L + D$

But the reflux ratio, R is:

$R = L/D$

$L = R\ D = 3 \times 337$

$= 1011$ lb mole/h

Mass balance in the rectifying section:

$V = L + D$

$= 1011 + 337$

$= 1348$ lb-mole/h

Component balance in the rectifying section of the column

At the feed stage, since 30 percent of the feed is vaporized, q = 0.7. L′ from the definition is:

In the stripping section of the column:

The component balance in the stripping section of the column

Slope of the q – line is:

From Equation 10 – 267 is:

where

Solve the material and component balances for the new conditions

$1000 = D + B$

$1000 \times 0.4 = D \times 0.99 + 0.05B$

$400 = 0.99D + 0.05B$

Solving these simultaneously gives B = 628 lb-mole/h and D = 372 lb-mole/h.

Recalculate S for the new conditions:

$= 7.54$

or

This gives the quadratic equation:

Solving for R gives:

= 3.58

Since D = 372 lb-mole/h, L = 3 × 372 = 1332 lb – mole/h

= 1332 + 0.7 × 1000 = 2032 lb-mole/h.

= 2032 – 628 = 1404 lb-mole/h

Therefore the computer control will therefore require increasing the boil-up rate from 1048 lb-mole/h to 1404 lb-mole/h.

Summary

Variable	This variable gives a measure of	lb-mole/h
L	Rectifying section liquid load	1011
V	Rectifying section vapor load Condenser duty	1348
L′	Stripping section liquid load	2033
V′	Stripping section vapor load Reboiler duty	1406

Example 10-25: Calculation of Minimum Number of Theoretical Trays/Plates/Stages at Total Reflux for a Binary System

A finishing column is required to produce trichlorethylene at 99.9% (vol.) purity from 10,000 lb/hour of a feed of 40% (wt.) trichlorethylene and 60% (wt.) perchlorethylene. Only 1 % (vol.) of the trichlorethylene can be accepted in the bottoms.

Because the process system that will receive vents from this condensing system is operating at 5 psig, allow 5 psi pressure drop to ensure positive venting and set top of tower pressure at 10 psig.

Feed (158°F)	(1)Wt %	(2)Mol Wt	(1) (2)Mols	MolFraction
Trichlorethylene	40	131.4	0.00304	0.456
Perchlorethylene	60	165.9	0.00362	0.544
Total	100		0.00666	1.000

Avg mol wt 1.00/0.00666 = 150.0

Overhead

Overhead temperature for essentially pure products at 10 psig = 223°F from vapor pressure curve.

Bottoms

Allow 10 psi tower pressure drop, this makes bottom pressure = 20 psig = 1,800 mm Hg.

Material balance:

$F = D + B$

Feed rate: mols/ hr =

$66.7 = D + B$

or

$B = 66.7 - D$

Component balance:

$F x_1 = D x_1 + B x_1$

$(66.7)(0.456) = D(0.999) + B(0.01)$

$30.4 = 0.999 D + 0.01 (66.7 - D)$

$D = 30.05$ mols/hr

bottoms: $B = F - D$

$B = 66.7 - 30.05 = 36.65$ mols/hr

Bottomscompositions	**Molfraction**	**Mols/hr**	**V.P. (316°F) mm Hg**	**V.P. mm Hg.**
Trichloroethylene	0.01	0.3665	4,200	42
Perchlorethylene	0.99	36.2835	1,780	1,762
	1.00	36.65		1,804 mm

The 1,804 mm compares to the balance value of 1,800 mm ≅ 20 psig.

Overhead Composition	Mol Fraction	Mols/hr
Trichlorethylene	0.999	30.02
Perchloroethylene	0.001	0.03
	1.000	

Relative Volatility: overhead conditions

Bottom conditions:

Thermal Condition of the Feed at 158°F

At conditions of feed tray, assume pressure is 15 psig ≅ 1,533 mm Hg. Determine bubble point:

Component	x_{iF}	Assume t = 266°F V.P mm Hg	Partial Press. = VP. × x_{iF} mm Hg
Trichlorethylene	0.456	2,350	1,072
Perchlorethylene	0.544	880	478
			1,550

This is sufficiently close to 1,533 mm for practical purposes; the actual temperature might be 265°F, although plotted data are probably not that accurate. Because the feed enters at 158°F and its bubble

point is 266°F, the feed is subcooled.

Heat to vaporize one mol of feed.

Component	x_{iF}	Latent Ht. @ 266°F Btu/mol	$(x_{iF})(L_v)$ Btu/mol	(°F) cP @ 158°F Btu/mol	(x_{iF})(cP)(266°F − 158°F) Btu/mol
Trichlorethylene	0.456	12,280	5,600	30.9	1,523
Perchlorethylene	0.544	14,600	7,950	36.4	2,180
			13,550		3,703

Minimum Number of Trays at Total Reflux

$x_{D1} = 0.999$

$x_{Dh} = 0.001$

$x_{B1} = 0.01$

$x_{BH} = 0.99$

$\alpha_{avg} = 2.8$

For a total condenser system:

$N_{min} + 1 = 11.17$

$N_{min} = 10.17$ trays, not including reboiler

Summary

Minimum total physical trays in the column	= 10.17
Reboiler	1.0
For conservative design, add feed tray	1.0
Minimum number of total theoretical stages	12.17, say 12

Minimum Reflux Ratio

Because this is not feed at its boiling point, but subcooled liquid, the convenient charts cannot be used with accuracy. Using Underwood's general case:

(10-269)

Solve first for y_{1F}, assuming that the system follows the ideal (as it does closely in this instance).

(10-48)

This replaces drawing the equilibrium curve and solving graphically, and is only necessary since the “q” is not 1.0 or zero.

The α should be for the feed tray. However, the value of α = 2.8 should be accepted for feed tray conditions (not 158°F). It would not

be if this were predominantly a rectifying or a stripping operation.

Now, substituting to solve for $(L/D)_{min}$

Solving this quadratic:

$(L/D)_{min} = 0.644$

Reading Figure 10-40 for $(L/D)_{min}$ assuming a liquid feed at the boiling point, $(L/D)_{min} = 1.2$. This demonstrates the value of taking the thermal condition of the feed into account.

Actually, any point on one of the curves represents a condition of reflux and number of trays that will perform the required separation.

10.32. Number of Theoretical Trays at Actual Reflux

Assume actual reflux ratios of 1.2, 1.8, 2.25, 3.0 times the minimum and plot the effect on the number of theoretical plates using Gilliland plot.

Actual Reflux Ratio		(From Fig. 10-49a) $N-N_{min}$	N (Theo.)	Conserv Add 1 Feed, T N
0.772	0.0722	0.552	26.2	27

1.16	0.239	0.416	19.8	21
1.45	0.329	0.356	17.9	19
1.93	0.439	0.288	16.1	17

Read value from curve Figure 10-49a.

, so

N = 26.2

Note that these values are for theoretical trays and, as in earlier calculations, require conversion into actual number of trays. Efficiencies are generally in the range of 50–60% for systems of this type. This indicates a column of actual number of trays almost twice the theoretical number at the operating reflux.

Figure 10-53 presents the usual determination of optimum or near optimum theoretical trays at actual reflux based on performance. It is not necessarily the point of least cost for all operating costs, fabrication costs or types of trays. A cost study should be made to determine the merits of moving to one side or other of the so-called optimum point. From the Figure 10-53:

Figure 10-53

Relationship of reflux ratio and theoretical trays, for Example 10-25.

First choice actual reflux ratio, L/D = 1.33

Corresponding theoretical trays or stages, N = 18.6

Note that the 18.6 includes the reboiler, so physical trays in column = 17.6. Do not round-off decimal or fractions of trays until after efficiency has been included.

Tray Efficiency

Base at average column temperature of (158 + 266)/2 = 212°F.

Component	x_{iF}	Viscosity, cP	cP
Trichlorethylene	0.456	0.27	0.123
Perchlorethylene	0.544	0.36	0.196
			0.319

From Figure 10-54:

Figure 10-54

Empirical correlations of overall efficiencies for fractionation and absorption.

Efficiency = 47.5%

Actual Trays at Actual Reflux

Actual L/D = 1.33

Actual trays = 18.6/0.475 = 39.2 (including reboiler)

Physical trays: 39.2 − 1 (reboiler) + 1 (conservative, feed) = 39.2

Round-off to: 40 trays plus reboiler plus total condenser.

Note: If there is any reason to know that the efficiency of this system is usually lower (or in same chemical family), then either the efficiency should be reduced to account for this or extra trays should be added. In practice, the installed column might contain 40 trays to 50 trays [271].

Types of Tray

Tray details will be considered in a later example.

10.33. Estimating Tray Efficiency in a Distillation Column

Several empirical efficiency correlations have been developed from commercial equipment and some laboratory data, and serve most design problems for the average hydrocarbon and chemical system. They are empirical correlations, and hence application in new systems is unpredictable. For this reason, results for efficiencies should be evaluated by several methods to obtain some idea of the possible spread. In light of the American Institute of Chemical Engineers (AIChE) study discussed below, some of these methods can be off the range by 15−50%. Comparisons indicate these deviations are usually on the safe or low side. The relation of Drickamer and Bradford [16], in Figure 10-54, has been found to give good agreement for hydrocarbons, chlorinated hydrocarbons, glycols, glycerine and related compounds, and some rich hydrocarbon absorbers and strippers.

The relation of O'Connell [49] (Figure 10-54) has generally also given good results for the same systems, but generally the values are high. The absorber correlation of O'Connell (Figure 10-54) can be used as long as it generally gives lower values than the other two relations. It can be used for stripping of gases from rich oils provided great caution is exercised.

The area of absorption and stripping is difficult to correlate for the wide range of peculiarities encountered in such systems. The correlation of Gautreaux and O'Connell [22] allows a qualitative handling of tray mixing to be considered with overall and local efficiencies. In general, it agrees with the Drickamer correlation–at least for towers upto seven feet in diameter. Although the effect of liquid path must be apparently considered, the wide variety of tray and cap designs makes this generally possible, and the overall correlations appear to serve adequately.

The American Institute of Chemical Engineer's Distillation Tray Efficiency Research [2] program has produced a more detailed method than the shortcut methods, and correspondingly is believed to produce reliable results. This method produces information on tray efficiencies of new systems without experimental data. At present, there is not enough experience with using this method and its results to evaluate its complete range of application.

Murphree [85] developed “point” and “overall” distillation tray efficiencies, which are examined in detail in Reference 2. The expressions are [59]:

(10-270)

The plate/tray efficiency is the integrated effect of all the point

efficiencies.

(10-271)

(10-272)

where

y_i = average composition of vapor entering tray

y_o = average composition of vapor leaving tray

= composition of vapor in equilibrium with liquid flowing to plate below

= vapor composition entering local region

= vapor composition leaving local region

y_e = vapor composition in equilibrium with the liquid in the local region

The proposal for calculating column vapor plate efficiencies by MacFarland, Sigmund, and Van Winkle [86] correlates with the Murphree vapor plate efficiencies in percent:

(10-273)

where

y_n = average light key mol fraction of vapor leaving plate n

y_{n+1} = average light key mol fraction of vapor entering plate n

y_e = light key mol fraction of vapor in perfect equilibrium with liquid leaving plate n

Data from bubble cap and perforated tray columns for the Murphree vapor plate efficiencies are correlated [86]:

(10-274a)

or,

(10-274b)

Referenced to 806 data points for binary systems, Equation 10-274a gives absolute deviation of 13.2%, which is about as accurate, or perhaps more so, than other efficiency equations. Equation 10-274b uses the same data and has an absolute average deviation of 10.6%. See Example 10-26 for identification of dimensionless groups.

Example 10-26: Estimating Distillation Tray Efficiency by Equations (10-274a) and (10-274b) (used by permission of McFarland et al. [86])

Solving the problem defined in the following table will show the equations for estimating system physical properties and their relation to the calculation of Murphree vapor plate efficiencies:

System properties*	**Acetone**	**Benzene**
Molecular weight, M, lb/lb mole	58.08	78.11
Viscosity, μ lb/hr-ft	0.5082	0.8155
Parachor, [P]	162.1	207.1
API specific gravity coefficient [240]		
A	0.8726	0.9485
B	0.00053	0.00053

C	21.6	18.0
E	536.0	620.6
Operating data		
Acetone mole fraction, x_1	=	0.637
Benzene mole fraction, x_2	=	0.363
Temperature, T, °F	=	166
Superficial vapor mass velocity, G, lb/hr-sq ft	=	3,820
Vapor velocity, U_v, ft/hr	=	24,096
Weir height, h_w, ft	=	0.2082
Fraction free area, FA	=	0.063
*Used by permission of McFarland *et al.* [86].		

Liquid densities for pure hydrocarbon are calculated [240] as a function of temperature using the following equation for specific gravity:

(10-275)

The liquid density is then:

$\rho_L = (62.32)\ (s_{gL})$

For acetone,

$\rho_{L,1} = (62.32)\ [0.8726 - 0.00053\ (166) - 21.6/\ (536.0 - 166)] = 45.3$ lb/ft^3

For benzene,

$\rho_{L,2} = (62.32)\ [0.9485 - 0.00053\ (166) - 18.0/(620.6 - 166)] = 51.2$

lb/ft^3

Vapor densities are calculated from the ideal gas relation:

$\rho_V = MP_t/555(T + 460)$

where total pressure P_t is given in millimeters of mercury.

Mixture densities of the binary mixtures require knowledge of volume fraction for each component. The component molar volume is:

(10-277)

For acetone and benzene, respectively:

$V_{L,1} = (58.08)/45.3 = 1.282\ ft^3/lb\ mole$

$V_{L,2} = (78.11)/1.51.2 = 1.526\ ft^3/lb\ mole$

For the liquid mixture:

$M_{L,MIX} = x_1V_{L,1} + x_2V_{L,2} = (0.637)\ (1.282) + (0.363)\ (1.526) = 1.371\ ft^3/lb\ mole$

Then the volume fraction of a component is calculated assuming an ideal mixture.

(10-278)

For acetone and benzene, respectively:

$\upsilon_{L,1} = 0.817/1.371 = 0.596$

$\upsilon_{L,2} = 0.554/1.371 = 0.404$

The *liquid density* of the binary mixture is then:

(10-279)

$= (0.596)\ (45.3) + (0.404)\ (51.2)$

$= 47.6\ \text{lb/ft}^3$

The *vapor density* can be found in an analogous manner.

(10-280)

However, the example problem does not require a calculation for vapor density. Instead, the superficial vapor mass velocity G can be substituted into Equation 10-288 because:

(10-281)

Liquid viscosity of the binary mixture, when not reported with the experimental efficiency results, is estimated using:

(10-282)

The pure component viscosities are given in the literature [240, 241]

as a function of temperature.

For example,

$\mu_{L,MIX} = [(0.637)\ (0.5082)1/3 + (0.363)\ (0.8155)^{1/3}]^3$

= 0.609 lb/hr-ft

Liquid surface tension is calculated using the Sugden Parachor method [242]. Neglecting vapor density, surface tension for the liquid mixture is:

(10-283)

where σ is in dynes/cm, ρ is in gm/cm^3 and the parachor,

(10-284)

Values of the parachor are given in the literature [240].

Then the example gives:

$M_{mix} = (0.637)\ (58.08) + (0.363)\ (78.11)$

= 65.35 lb/lb mole

$[P]_{min} = (0.637)\ (162.1) + (0.363)\ (207.1)$

= 1 78.4

ρ_{min} = 47.6/62.32 = 0.7638 gm/cu cm

$\sigma_{min} = [(0.7638/65.35)\ (178.4)]^4$

$= 18.96$ dynes/cm

Diffusivity of the liquid light key component is calculated by the dilute solution equation of Wilke-Chang [243].

(10-285)

Wilke-Chang reported the recommended values for Ψ as follows: water, 2.6; benzene, heptane and ether, 1.0; methanol, 1.9; ethanol, 1.5; unassociated solvents, 1.0. The mixture parameter for the example problem is considered unity.

Then,

$D_{LK} = (3.24 \times 10^{-8})\ (65.35)^{1/2}\ (166 + 460)/(0.609)\ (1.282)^{0.6}$

$= 2.32 \times 10^{-0.4}\ ft^2/hr$

Dimensionless groups for the example problem are:

(10-286)

(10-287)

(10-288)

Murphree vapor plate efficiency is calculated two ways:

(10-274a)

(10-274b)

In this example, Equation 10-274b gives a more conservative design basis, where

A, B, C, E = constants in equation
D = molecular diffusion coefficient, sq ft/hr
E_M = Murphree vapor plate efficiency, %
FA = fractional free area
h_w = weir height, inches
G = superficial mass vapor velocity based on the crosssectional area of the column, lb/hr-sq ft
M = molecular weight, lb/lb mole
N = dimensionless number
P = pressure, consistent units
[P] = Sugden parachor
sg = specific gravity
T = temperature, °F
U = superficial velocity, ft/hr
V = molar volume, ft^3/lb mole
υ = volume fraction
x = mole fraction in the liquid
y = mole fraction in the, vapor
μ = liquid viscosity, lb/hr-ft
ρ = density, lb/ft^3
σ = surface tension, dynes/cm

Ψ = mixture parameter

Subscripts

i = component
L = liquid
LK = liquid light key
mix = binary mixture
n = plate number
t = total
V = vapor

Biddulph [90] emphasizes the importance of using point efficiencies rather than tray or overall column efficiencies, due to the wide fluctuations that often exist.

Kessler and Wankat [101] have examined several column performance parameters, and for O'Connell's [49) data presented in Figure 10–54, they propose equations that reportedly fit the data within about ±10%:

A. Distillation Trays

(10-289)

B. Plate Absorbers (data fit ±5%)

(10-290)

where

E_o = overall efficiency
H' = Henry's law constant, lb mole/ (atm) (ft^3)

P = pressure, atmospheres
α = relative volatility
μ = viscosity, centipoise, cP

Gerster [176] presents the results of studies on the tray efficiencies of both tray and packing contacting devices. Note that Gerster compares his work to the AIChE Manual [2].
In terms of the change in gas composition [2]:

(10-291)

where

E_G = overall column efficiency

E_{OG} = overall point efficiency in vapor terms (see Ref. 2)

y_{n+1} = component mol fraction in the gas to the point considered

y = component mol fraction in the gas from the point considered

y^* = composition the leaving gas would have if it left the point in equilibrium with the liquid

In Table 10-27 Proctor [178] compares efficiencies of sieve and bubble cap trays (plates). He concludes that the sieve design provides a 15% improvement in plate efficiency. To fully evaluate actual efficiencies in any particular system, the physical properties, mechanical details of the trays, and flow rates must all be considered. See also Reference 2.

Table 10-27 Comparative Effectiveness of Sieve and Bubble-Cap Trays/Plates [178]

Plate type	Vapor throughput, lb Mole/hr of dry H_2S	Overall plate efficiency, %	
		Cold tower	Hot tower
Sieve	18,200	69 ± 5	75 ± (8) *
Bubblecap	16,200	60 ± 5	69 ± 5

*See the discussion of accuracy of the plate efficiency results in the text. Used by permission of the American Institute of Chemical Engineers; all rights reserved.

Strand [179] proposes a better agreement between experimental and predicted efficiencies when recognizing a liquid bypassing factor to correct predicted values determined by the AIChE method. The results suggest that, for the representative systems studied, recognition of a liquid bypassing factor for a tray can increase the agreement of the AIChE method results by as much as 5 or 10%. A vapor bypassing effect was not required to correlate the data. Because the Murphree vapor efficiencies vary considerably for various systems, the data in Reference 179 can only be a guide for systems not studied.

This suggests that caution must be exercised when establishing a tray efficiency for any type of contacting device by: (1) using actual test data if available for some similar system; (2) comparing several methods of predicting efficiency; and (3) possibly using a more conservative efficiency than calculated to avoid the possibility of ending up with a complete column with too few actual trays–a disastrous situation if not discovered prior to startup.

Sakato [180] evaluates the degree of mixing of the liquid as it flows across a tray and its effect on the tray efficiency, Figure 10-55. For plug flow the liquid flows across the tray with no mixing, while for partial or “spot” mixing as it flows over the tray, improved tray

efficiency can be expected. For a completely mixed tray liquid, the point efficiency for a small element of the tray, E_{OG}, and tray efficiency, E_{MV}, are equal.

Figure 10-55

Effect of vapor mixing on tray efficiency.

Reprinted by permission, Sakato, M., The American Institute of Chemical Engineers, Chem. Eng. Prog., V.62, No. 11 (1966), p. 98, all rights reserved; reprinted by permission from Lewis, W. K., Jr., Ind. & Eng., Chem. V. 28 (1936), and by special permission from Fractionation Research, Inc.

From Figure 10-56 the effect of mixed and unmixed "pools" of liquid can be noted. For a completely mixed tray, there is no concentration gradient from inlet to outlet, and therefore the entire tray has a uniform composition. The degree of mixing across the tray, as determined by the number of discrete mixing pools on it, has an effect on the relationship between E_{OG} and E_{MV} as a function of λ, where:

$\lambda = mV/L$
m = slope of vapor-liquid equilibrium curve
V = vapor rate, lb mols/hr
L = liquid rate, lb mols/hr

Figure 10-56

Typical effect of liquid mixing on tray efficiency.

Reprinted by permission, Sakato, M., The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 62, No. 11 (1966), p. 98, all rights reserved; reprinted by permission from Lewis, W. K., Jr., Ind. & Eng., Chem. V. 28 (1936), and by special permission from Fractionation Research, Inc.

Hughmark [181] has proposed empirical correlations for better fit of experimental data to transfer units and thus tray efficiency comparison with the AIChE method [2].

10.34. Batch Distillation

Specialty chemicals represent some 15% of worldwide chemical production, and is worth approximately $15 billion annually. Batch distillation is the most suitable unit operation for the separation such material [278]. This is because of its flexibility and versatility in separating small amounts of mixtures, in order to obtain products which are expensive, and to recover compounds or solvents that could adversely affect ecosystems if lost. This flexibility also allows uncertainties in feed stock or product specification to be dealt with, once several mixtures can be handled by switching column operating conditions. Batch distillation is therefore preferable to continuous distillation when small quantities of high technology/high value added chemicals and biochemicals need to be separated.

Previously, calculations for batch processes were based on trial and error; however, such design may not result in an economic optimum. Studies have been conducted with a view of understanding and developing improved design methods for batch processing [278, 279, 280, 281].

In batch distillation, an initial quantity of materials is fed to the still, and during the separation, one or more component is collected in a receiver. The process results in a change of composition in the still over time. Two modes of operation are possible:

1. Maintain a constant top product specification, the reflux ratio and hence increase continuously.
2. If the reflux ratio remains constant throughout the operation, the quality of the top product will decrease with time. To obtain a product specification of composition x_D, the initial product removed will have a composition x greater than x_D and the process will stop when x is less than x_D such that the average composition over the whole operation is equal to x_D.

In these modes of operation, the McCabeThiele construction may be used for calculations–as illustrated for continuous distillation process.

Batch distillation [129, 130, 131, 133, 138, 140, 142, 170] may be preferred to continuous distillation for process requirements in which: (1) feed composition may change from batch to batch; (2) batches are relatively small fixed volumes of a mixture wherein certain components are to be separated into relatively pure components, leaving a heavier residue; (3) the process improvement requirement is on an irregular cycle; and (4) there is negligible holdup in any column used and condenser relative to that in the receiver and kettle. The system operates on a fixed feed quantity, thereby yielding a fixed distillate and residue (see Figures 10-57 and 10-57a). In batch operation, the kettle is charged with a fixed amount of liquid, and the vapors; the rise either through a trayed or packed column contacting section above the "pot." This then condenses the vapors, and collects the components, separated according to their boiling points. Thus, the separation can be developed by the boilup to collect only, or nearly only, the light boilers, then the next boiling fraction, etc., until the light ends and the heavies or residues are at the collection and concentration levels desired.

Figure 10-57

Batch operations: constant reflux ratio and variable overhead composition for fixed number of theoretical stages/trays.

Used by permission, Treyball, R. E., Chem. Eng. Oct. 5(1970), p. 95.

Batch distillation is versatile, since the same unit may be used to process different feeds and generate multiple products. A typical run may be from a few hours to several days depending on the process. The approach is used for purifying materials of high value that are generally handled in relatively small quantities. The techniques are used extensively in the Fine Chemical and Pharmaceutical industries to achieve a variety of processing objectives, such as:

- Removal of impurities/unwanted components from a reaction mixture.
- Solvent change between reaction stages on multistage syntheses.
- Solvent recovery.
- Water removal and drying.
- Heat removal control for reactions at reflux.
- Concentration prior to crystallization.

The simplest form of batch distillation involves a single separation stage (represented by the act of evaporation), where a large difference in volatility exists between the components to be separated. In such a case, there is no need for a fractionating column. Simple batch distillation (sometimes referred to as still or pot distillation) provides only one theoretical plate (Figure 10-58). When the difference in volatility between the components to be separated is relatively small,

or when operating over narrow composition ranges, a rectification section is added. Overhead facilities are also required to provide control of reflux ratio and layer of reflux when handling heterogeneous azeotropes. An alternative to a simple batch process is constant level batch distillation, where solvent is fed continually to the still batch to keep the liquid level constant. Figure 10-4a shows a typical multistage batch distillation, where a stage column is placed above the reboiler. Here, the reflux is returned to the column. There are instances where the distillate is withdrawn continually until the column is shut down and drained. Alternatively, where no distillate is withdrawn, the composition of liquid in the accumulator changes. When the distillate in the accumulator reaches the desired composition, both the accumulator and the reboiler are drained. Economic design and operation of batch distillation systems must consider for many time dependent factors. The complex mathematics of the process often involves using computer simulation tools such as BATHFRAC (Aspen Technology, 2000) and Chemcad-BATCH (Chemstations, Inc. 2000) and academic MultiBatch DS to solve batch problems.

Figure 10-58

Schematic diagram of a simple batch distillation.

10.34.1. Simple or Differential Distillation

In this process, a mixture of initially soluble liquids of different volatilities is heated to the bubble point to evolve vapor that is richer in the more volatile components. This vapor is removed as it is generated, so that the less volatile components accumulate in the

liquid. Thus, while the vapor formed at any instant is in equilibrium with the liquid in the still, the total amount of vapor collected over the entire distillation operation will not be in equilibrium with the final residual liquid remaining in the still. Figure 10-59 shows a schematic diagram of a simple batch distillation system. Here, the vapors are withdrawn from contact with the liquid and condensed as fast as they are formed.

Figure 10-59

Schematic diagram of a multistage batch distillation.

In a typical batch distillation, the still pot will be charged with feed F. For stills which must accommodate larger quantities of feed, the bottom portion of the still, is usually cylindrical or spherical in shape, and is larger than the column. After charging, the still bottom is heated until the desired reflux rate is established. Then, distillate is withdrawn from the overhead accumulator until the desired end point is reached.

In the cases where reflux is employed, either of two general types of control may be used. These are:

1. Constant reflux ratio (R = L/D) with varying distillate composition.
2. Constant overhead purity with variable reflux ratio.

Control type 1 can be accomplished with various types of stream splitting devices, and type 2 can be performed by placing reflux flow

rate in a control loop to maintain constant column top temperature. In either case, one or more distillate cuts may be produced after startup by collecting the distillate in one or more receivers as the process of distillation is carried out. At the end of the process, the heat supply is turned off and the remaining residue can be removed from the pot as a bottom product.

The following assumptions are made for all cases:

1. Hold-up of liquid and vapor in all parts of the still except the bottom (residue) is negligible.
2. Reflux, if any, is at its bubble point.
3. The still or bottom of the still constitutes one equilibrium stage.
4. The residue is completely mixed at all times during the distillation process.
5. The system is a binary one with constant relative volatility.
6. Heat loss from the column is negligible.
7. Equilibrium stages are numbered from the bottom to the top of the distillation column, starting with the first stage after the still bottom or pot.
8. Where the reflux is used, the liquid downflow rate (L mol/s) does not vary from tray to tray (i.e. constant molal overflow).

Equations

Consider a binary batch of W moles of liquid charged to a batch still. It contains a mixture of two liquid components A and B with A being more volatile (see Figure 10-59).

After a certain time t, a differential amount of liquid dW is vaporized.

Therefore, the amount of liquid left in the still pot is

(10-292)

The amount of A present in the initial charge

(10-293)

The amount of A in the vapor

(10-294)

Because of evaporation, the liquid composition will decrease from x to (x – dx) with respect to component A.

Therefore, the amount of A left in the still pot is:

(10-295)

Applying the material balance on component A

Amount of A initially present in the still pot =	Amount of A left in the still pot	Amount of A + vaporized

(10-296)

(10-297)

Neglecting the higher order (dx dW), Equation 10-297 becomes:

(10-298)

Rearranging Equation 10-298 gives:

(10-299)

Rearranging Equation (10-299) and integrating from the initial charge condition of W_o and to W and after distillation gives:

(10-300)

(10-301)

or

(10-301a)

where

x = composition of A in the binary liquid mixture, mole fraction

x_{wo} = initial composition of liquid in the still, mole fraction
x_w = final composition of liquid in the still, mole fraction
y = composition of vapor of A, mole fraction
W_o = moles of component A in the still pot before distillation
W = moles of component A after distillation

Equations (10-301) and (10-301a) are called the Rayleigh equations. Although the time, t does not appear explicitly in the derivation of Equation 10-301 or 10-301a, it is present implicitly since, W, w_x and x_{wo} are all time-dependent.

In a total condenser, where x and y are now in equilibrium, the equilibrium expression of Equation 10-301a can be integrated graphically, analytically or numerically by:

(10-302)

An alternative form of Equation 10-301 is as follows.

At any time during the course of batch distillation, there are L moles of liquid of composition x with respect to A in the still pot. If in time δt, an amount of dD moles of distillate of mole fraction y^* in equilibrium with the liquid is vaporized, then the following material balances apply:

Operation	Total material	Component A
Moles in	0	0
Moles out	dD	
Moles accumulated	dL	d (L x)
Accumulation = In	0 – dD = dL	

- Out		
	i.e. – dD = dL	*i.e.*

Substituting –dD = dL in the material balance for component A gives:

(10-303)

Rearranging Equation 10-303 gives:

(10-304)

Integrating Equation 10-304 between F, x_F, moles of feed and composition charge initially in the still respectively, and W moles of residual liquid of composition x_w after distillation is carried out for a certain period.

Equation 10-304 becomes:

(10-305)

or

(10-306)

The integrating procedure for Equation 10–302 or Equation 10–306

is:

1. Plot x–y equilibrium curve.
2. At a series of x values, find y – x.
3. Plot 1/ (y − x) .vs. x or fit it to an equation.
4. Graphically or numerically integrate from x_{wo} to w_x as shown in Figure 10–60

Figure 10-60

Graphical integration of Rayleigh's equation.

5. From Equation 10-301, find the final charge of material in the still pot.

(10-307)

(10-308)

6. The average distillate concentration of component A can be determined from the mass balances as follows: For a binary batch distillation, the mass balance around the entire system for the total batch operation is:

(10-309)

Component balance on A is:

(10-310)

Combining Equations (10-309) and (10-310), the average distillate concentration of component A is:

(10-311)

(10-312)

The area under the curve of 1/(y – x) vs. x can be integrated numerically by Simpson's rule (see Appendix H). If the ordinate in Figure 10–60 is called f(x), then the form of Simpson's 1/3 rule is:

(10-313)

where

h = step size

x_{in} = initial component, mole fraction

$x_{fin.}$ = final component, mole fraction

y^* = mole fraction in equilibrium with the liquid being vaporized.

Equation (10-301) is easily integrated only when pressure is constant, temperature change in the still is relatively small (close-boiling mixture) and K-values are composition dependent. Then y = Kx, where K is approximately constant, Equation (10-300) becomes:

(10-314)

For a binary mixture, where the relative volatility α is assumed constant over the concentration range then.

(10-48)

Substituting Equation (10-48) into Equation (10-300) gives:

(10-315)

(10-316)

where

D = instantaneous distillate rate, mol/h
F = moles of feed
W = moles of liquid left in still.
x_F = composition of liquid in the feed, mole fraction.
$x = x_w$ = composition of liquid left in still, mole fraction
$y = y_D = x_D$ = instantaneous distillate composition, mole fraction.

Example 10-27

A binary mixture of ethanol and water at 101.3 kPa (1 atm) is charged to a batch still. The initial charge is 100 kmol of ethanol-water mixture at 18 mole per cent and the mixture is to be reduced to a minimum ethanol concentration of 6 mol per cent. Determine the following:

1. The amount of ethanol in the leftover after distillation.
2. The amount of ethanol in the distillate.
3. The distillate composition.

Solution

From the vapor–liquid equilibrium data (VLE) (see Figure K-1b in appendix-K). Table 10-28 shows the equilibrium data of ethanol-water system shown. Plotting $1/(y^* - x)$ versus x and integrating between the limits of 0.06 and 0.18 for x, the area under the curve (Figure 10-61) is found, using the Excel spreadsheet (Example 10-27.xls), to be 0.358. Then, ln (F/W) = 0.358, from which W = 100/1.43 = 69.91 kmol. The liquid remaining in the still consists of (69.91)(0.06) = 4.19 kmol of ethanol and 65.72 kmol of water. By material balance, the total distillate must contain (18.0 – 4.19) = 13.81 kmol of ethanol and (82 − 65.72) = 16.28 kmol of water. The total distillate is 30 kmol, and the distillate composition is 13.81/30 = 0.459 mole fraction of ethanol.

Table 10-28 Vapor–Liquid Equilibrium Data of Ethanol-Water Mixture.

x		
0.18 = x_{in}	0.5169	0.3369

0.16	0.5022	0.3422
0.14	0.4853	0.3453
0.12	0.4642	0.3442
0.10	0.4379	0.3379
0.08	0.4052	0.3252
0.06 = x_{fin}	0.3529	0.2929

Figure 10-61a–b Snapshot of a batch distillation calculation for a binary ethanol-water system. **c–d** Snapshot of a batch distillation calculation for a binary ethanol-water system.

An Excel spreadsheet (Example 10-27.xls) has been developed for Example 10-27, and snapshots of the calculation are shown in Figures 10-61a–d.

A summary of the Excel spreadsheet calculations is shown in the Table 10-29.

Table 10-29 A Summary of the Results of Example 10-27.

Composition	Feed, kmol	Residue, kmol	Distillate, kmol
Ethanol	18	4.19	13.81
Water	82	65.72	16.28
Total	100	69.91	30.09

10.34.2. Differential Distillation; Simple Batch, No Trays or Packing; Binary Mixtures, No Reflux

For systems of high (above approximately 3.0) constant relative volatility the Raleigh equation can be expressed:

(10-317)

or

(10-318)

The above Equation 10-318 requires graphical integration or numerical using Simpson's rule, where

B_{T0} = total moles of liquid in bottom of still at start, T_0

B_{T1} = total moles of liquid in bottom of still at time, T_1

x_0 = mol fraction of component, i, in bottoms B_{T0} at start, time T_0

x_l = mol fraction of component, i, bottoms B_T, at time T

α = relative volatility of light to heavy components

y^* = equilibrium value of x_i

The condensed vapor is removed as fast as it is formed.

The results of either relation allow the plotting of an instantaneous vapor composition for given percentages of material taken overhead.

The outline of Teller [70, 133] suggests using the differential form above. The vapor is assumed to be in equilibrium with the liquid.

1. Calculate or obtain an x-y equilibrium diagram for the light

component.

2. Select values of x_i and read equilibrium values of y_i from Step 1 above.

3. Calculate values of $1/(y_i - x_i)$ and tabulate.

4. Plot curve of $1/(y_i - x_i)$ versus x_i; see Figure 10-60, graphical integration by Simpson's rule.

5. From the plot of Step 4, determine the area under the curve from initial bottoms concentration of x_{i0} mol fraction at beginning of distillation down to the final lower concentration of x_{i1} in bottoms.

6. The area from Step 5 represents

where

W_{i1} = the final kettle/still pot content, mols

W_{i0} = the initial kettle/still pot content, mols

or, for constant relative volatility for a binary mixture for a simple still/kettle/bottoms pot with no internal packing or trays, a direct analytical solution is [133]:

(10-319)

where

W_1 = content of kettle at any time, mols

W_{io} = initial content of kettle, mols

W_o = mols liquid initially in still or kettle

α = relative volatility

D = distillate rate, mols/h

L = liquid rate, mols/h

V = vapor rate, mols/h
x = mol fraction of a specific component in liquid
y = mol fraction of a specific component in vapor
θ = time, hours

Subscripts:

D = distillate related
i, or o = initial
l = final or later
w = relating to bottoms (kettle/still pot)

7. For each value of x, and the values of (B_{To}/B_{TI}) found above, calculate (100), the percent of material taken overhead.
8. A plot of the distillate composition, y versus percent distilled (from Step 7) will show the value of the instantaneous vapor composition.

The usual form of the Raleigh equation [130] is for the conditions of a binary simple differential distillation (no trays or packing), no reflux, but with constant boilup.

(10-320)

For a binary mixture, the values of x and y can be obtained from the equilibrium curve. Select values of x_l and read the corresponding value of y from the equilibrium curve. Tabulate values of $1/(y - x)$, and plot versus x_l, resulting in a graphical integration of the function $dx/(y - x)$ [130] between x_o and x_l. This system would have no column internals and no reflux.

10.34.3. Simple Batch Distillation: Constant α, with Trays or Packing, Constant Boilup, and with Reflux [129] Using x-y Diagram

The system material balance, from Treybal [129], using a heated kettle and distillation column following a McCabeThiele diagram, using reflux, but having only a batch (kettle) charge:

(10-321)

(10-322)

(10-323)

G = boilup overhead, mol/hr
L = mols reflux in the column
D = overhead receiver contents, mols

Starting with an empty overhead receiver, the time θ_1 to condense D mols of vapor to fill the receiver, when the vapor boilup rate is G mols/hr.

θ_1 = D/G during which time the receiver is filling and there is no reflux and the kettle mixture follows a Raleigh distillation [129]. Under this condition, when the distillate receiver just becomes full, let the composition of the kettle contents be x_{Si}, then

(10-324a)

or

(10-324b)

Equation (10-324b) may be solved for x_{Si} by trial and error.

After this reflux runs down the column the desired lighter components leave, and a desired residual composition is left, following the Raleigh equation to express the material balance.

Most batch distillations/separations are assumed to follow the constant relative volatility vapor–liquid equilibrium curve of:

(10-48)

After filling the receiver, reflux runs down the column at the same molar rate as the vapor back up (L = G). The operating line has a slope of 1.0. Then there are “n” plates/trays between composition x_p and x_l (the mol fraction in distillate). As the distillation continues, the operating line moves closer to the 45° line of the diagram, and x_l and x_p (and x_S) become richer and leaner, respectively, until at the end x_l becomes x_D and x_S becomes x_W. The required time is θ_2 hours.

During a batch distillation at constant pressure, the temperature rises to accomplish the separation as the more volatile component's concentration is reduced in the bottoms (kettle) or residue.

For a batch differential distillation where no reflux is used, there is

only boilup of a mixture of the desired lighter component, which leaves the kettle, and a desired residual bottoms composition is left in the kettle. This type of distillation follows the Raleigh equation for expression of the material balance. However, while simple, not having tower packing or trays or reflux is not useful in many industrial applications due to the low purities and low yields involved. Repeated charges of the distillate back to the kettle and redistilling will improve overhead purity.

The minimum number of plates required will occur when the operating line coincides with the 45° diagonal [129]. But this corresponds for an infinite time for separation:

(10-325)

For an operating line with a slope of unity, from Smoker's equation:

(10-326)

Equilibrium curve:

(10-48)

Operating line: $y = x + b$ (see Reference 129 for diagram). They intersect at $x = k$.

Then,

(10-327)

where

$b = (\alpha k/c) - k$

$c = 1 + (\alpha - l)k$

Then,

(10-328)

Coordinates:

$x_1' = x_1 - k$

$y' = y - (b + k)$

$x_p' = x_p - k$

For the more volatile component at any time:

(10-329)

$b = y_S - x_p$

(10-330)

10.34.4. Fixed Number Theoretical Trays: Constant Reflux Ratio and Variable Overhead Compositions

Raleigh equation from [130]:

(10-331)

where

W_o = mols liquid mixture originally charged to still

W_1 = mols final content in still

x_{Wo} = initial mol fraction of more volatile component in mixture

x_W = composition of liquid in still, mol fraction

x_i = mol fraction of component in liquid phase

x = mol fraction of more volatile component in liquid

x_D = instantaneous mol fraction of the component in the distillate that is leaving the condenser at time θ.

x_{Di} = initial distillate composition, mol fraction

x_i = mol fraction component of i in liquid phase

y_i = mol fraction of component of i in the vapor phase

D = mols of distillate per unit time, or mols of distillate at time θ, or distillate drawoff

K_A, K_B = equilibrium vaporization constants for A and B, respectively

L = mols of liquid per unit time, or liquid return to column

P = distillate drawoff percentage = 100/ (R + 1)

P_i = pure component vapor pressure, mm Hg

R = reflux ratio (liquid returned to column)/(distillate drawoff); subscripts indicate number of plates, R_{min}

V = vapor rate up column, mols per unit time

θ = time

10.34.5. Batch Distillation with Constant Reflux Ratio, Fixed Number of Theoretical Plates in Column, Overhead Composition Varies

In this mode of operation, the column has a fixed number of theoretical trays that operates at constant reflux ratio. The reflux is set to a predetermined value, maintained throughout the run. As long as

the still composition is changing in the batch, so the composition of the overhead product (i.e. distillate) continuously changes. The column holdup is assumed to be negligible, and progress is illustrated in Figure 10-62. The figure shows the profiles of the compositions in the distillate and reboiler with respect to total mols distilled. The mole fraction of the lighter component in the reboiler (still-pot) falls off steadily and continuously with the increase of total mols distilled (i.e. as long as the distillation process continues). The profiles of the curves are strongly influenced by factors such as the degree of relative volatility, the predetermined value of the reflux ratio and the number of theoretical plates. Distillation is continued until the average distillate composition attains the desired value. The overhead product is then diverted to another receiver and an intermediate cut is withdrawn until the leftover liquor in the still-pot meets the required specification. This intermediate cut is normally added to the next batch.

Figure 10-62

A typical example in a binary batch distillation at a constant reflux ratio showing the variation in distillate and reboiler compositions with the amount distilled.

At any time θ [131]:

(10-332)

where

S_o = mols initially charged to still
S = mols in mixture in still at time θ
x_D = instantaneous mol fraction of a more volatile component in the distillate entering the receiver at time θ
x_{so} = mol fraction of a more volatile component in the initial still charge
x_s = mol fraction of a more volatile component in the still after time θ
D = mols of distillate at time θ

To solve the right side of the Raleigh-like equation, integrate graphically by plotting or numerically:

The area under the curve between x_{so} and x_s is the value of the integral. Plot the equilibrium curve for the more volatile component on x − y diagram as shown in Figure 10-63. Then, select values of x_D from the operating line having the constant slope, L/V. From equation

Figure 10-63

Variable reflux ratio at various theoretical plates to achieve a specified separation from x_0 kettle to x_3 distillate overhead. Note, all reflux ratios shown yield same separation, but require different numbers of theoretical places. D = Distillate; F = Kettle conditions x_0, y_0 at equilibrium.

$[L_{n+l} = L_n + D$

are drawn from the intersection of x_D and the diagonal. Then from these L/V lines, draw steps to the equilibrium curve, the same for a binary McCabeThiele diagram [130]. The proper operating line is the one that requires the specified number of theoretical plates (stages) in moving stepwise down from the initial desired distillate composition to the composition of the mixture initially charged to the kettle (or pot or still). The kettle acts like and is counted as one theoretical stage or plate. The intersection of the last horizontal step (going down the column) from x_D with the equilibrium curve is the still or kettle bottoms composition, x_W, at the completion of this batch distillation. Using the system material balance and the constant reflux ratio used (L/V), calculate the total vapor generated by the kettle. The heat duty of the kettle can be calculated using the appropriate latent and sensible heat of the mixture components.

For a constant reflux ratio, the value can be almost any ratio; however, this ratio affects the number of theoretical plates and, consequently, actual trays installed in the rectification section to achieve the desired separation. Control of batch distillation is examined in Reference 134.

The internal reflux ratio is L/V, and is the slope of the operating line. The external reflux is [133]:

$R = L/D$

and

$V = L + D$

$V/L = L/L + D/L = 1 + l/R = (R + l)/R$

Then

$L/V = R/(R + 1)$

and

$R = (L/V)/[1 - (L/V)]$, see Figure 10-63.

10.34.6. Minimum Reflux Ratio

Point F on Figure 10-63 represents conditions in the kettle or still with x_i, y_i, or x_o, y_o. Line DF represents slope of the operating line at the minimum reflux ratio. The stepwise development from point D cannot cross the intersection, F, where the slope intersects the equilibrium line, and leads to an infinite condition, as point F is approached. Thus, an infinite number of theoretical trays/stages is required, and thus a minimum reflux requirement. Minimum reflux is calculated [133]:

(10-333)

Note that $y_D = x_D$ on the diagonal of the equilibrium plot, and y_i and x_i are points of intersection with the equilibrium curve. For an abnormal equilibrium curve (as compared to regular or normal shape) see Figure 10-64.

Figure 10-64

Minimum reflux for abnormal equilibrium curve for batch operation, constant reflux ratio.

Once the minimum reflux ratio has been established (which is not an operating condition), then a practical reflux ratio from 1.5 – 10 times the minimum can be selected. The larger the reflux value down the column, the more vapor has to be boiled up, and the greater will be the required column diameter; the column height will however be less. An economic balance must therefore be established.

Solving the typical Raleigh equation:

(10-334)

Calculating W from Equation 10-334, an overall component balance gives the average distillate composition.

(10-335)

The average distillate composition will be:

(10-335a)

A trial-and-error calculation is necessary to solve for W until a value is found from the ln (W_i/W) equation above that matches the x_{Davg} which represents the required overhead distillate composition. By material balance:

$V = L + D$, and $R = L/D$

$V/D = L/D + 1 = R + 1$

$D = V/(R+1)$

(10-336)

(10-337)

Then from Equations (10-334) and (10-336):

(10-338)

(10-339)

An alternative equation in determining the time (hours) for distillation is:

(10-340)

Referring to Figure 10-65, the constant internal reflux ratio, L/V is shown for several selected values of reflux ratios [131]. Only one can be used at a time for actual operation. Starting at the intersection of the diagonal line (distillate composition), step off the theoretical plates. For example, from Figure 10-65 at constant reflux, using operating line No.1, starting at $x_D = 0.95$, for one theoretical plate, the bottoms composition in component A would be approximately $x_w = 0.885$; then going down one more plate at the same L/V, a second

theoretical plate yields a bottoms of x_B or x_w = 0.83, still yielding x_D = 0.95. If the L/V for the operating line No.4 is used (same slope as line No.1), then the expected performance would be x_D = 0.60, and after one theoretical plate, the bottoms would be 0.41 at the same reflux ratio as the first case; and for x_D = 0.60 and two theoretical plates, x_w = 0.31, where

D = Distillate rate, moles/hr.
L = Liquid flow rate, moles/hr.
V = Vapor flow rate, moles/hr.
x = Mole fraction of substance in liquid.
y = Mole fraction of substance in vapor
W = Bottom or pot liquors
θ = Time, hrs.

Figure 10-65

Batch distillation: constant reflux ratio after McCabeThiele diagram.

Revised/adapted and used by permission, Schweitzer, P. A., Handbook of Separation Techniques for Chemical Engineers, McGraw-Hill Book Co. (1979); also reprinted by special permission, Chem. Eng. Jan. 23 (1961), p. 134, © 1961, by McGraw-Hill, Inc. New York.

Subscript

D = Distillate
i = Initial condition
w = Pot liquors.

Calculation Procedure

1. Plot the vapor–liquid equilibrium (VLE) diagram of the lighter component of the mixture on the x–y coordinate.
2. Draw the 45° line (x = y line).
3. Select several values of x_D and draw operating lines having the same slope (L/V = constant) through the point of intersection of x_D and the 45° line (Figure 10-65).
4. Draw the steps between the operating line and equilibrium curve, as in the McCabeThiele method. This technique simultaneously solves the operating line and equilibrium equations.
5. Select the correct line. It is the line that requires the specified number of theoretical stages in going from the initial distillation composition to the composition of the mixture initially charged to the still.
6. Count the still-pot as one theoretical stage.
7. The intersection of the last horizontal step (i.e. moving down from x_D) with the vapor-liquid equilibrium curve is the composition x_w of the liquid in the still.

If the internal reflux ratio (L/V) is specified, the initial distillate composition, x_D represents the composition of the first drop of distillate. In such cases, the usual procedure for determining the initial distillate composition is trial and error. Once x_D is determined, successively lower distillate compositions are arbitrarily chosen and tabulated beside the corresponding composition of the liquid mix in the still. When the still concentration is reduced to the desired value, the distillation is stopped.

Example 10-28

A batch containing 500 kmol of an ethanol-water mixture containing 50 mol % ethanol is to be distilled at 1.013 bara. The rate of vaporization is 80 kmol/h and the product specification is 80 mol %

ethanol having a liquid vapor ratio (L/V) of 0.78. If the reflux ratio is 3.0, and the system has 6 theoretical plates including the pot. Calculate the following:

1. The kmol of product that will be obtained.
2. The average or accumulated distillate composition.
3. The time for distillation process.

Solution

Figure 10-66 is a schematic diagram of a multistage batch distillation.

Figure 10-66

Multistage batch distillation.

Example 10-28 is solved by trial and error, *i.e.*

1. Select several arbitrary values of x_D.
2. Locate them on the 45° line.
3. From each point, draw an operating line at a slope of 0.78
4. Step off 6-steps on each such line.
5. At the end of the 6th step, find the equilibrium value of the bottoms.
6. Integrate

between $x_i = 0.5$ and various lower limits of x_W (bottoms).

7. Determine W from:

8. Calculate the average distillation composition.

(10-335)

9. Check whether the calculated value of x_{Davg} is the same as (or approximates to) the required product specification.

Using the vapor–liquid equilibrium diagram of the ethanol-water system, plot an x-y diagram, and draw a number of operating lines at varying x_D values on the 45°– line with a slope (L/V) of 0.78. Step-off six plates on each to find the equilibrium value of the bottoms product. Figure 10-66Figure 10-66 and Figure 10-66 show McCabeThiele plots at varying x_D with slope L/V = 0.78, and Table 10-30 shows the results of these plots.

Figure 10-66

(a). Graphical estimation of liquid composition at x_D = 0.80 and L/V = 0.78 in Example 10-28. (b). Graphical estimation of liquid composition at x_D = 0.795 and L/V = 0.78 in Example 10-28.

Figure 10-66

(c). Graphical estimation of liquid composition at x_D = 0.79 and L/V = 0.78 in Example 10-28. (d). Graphical estimation of liquid composition at x_D = 0.785 and L/V = 0.78 in Example 10-28.

Figure 10-66

(e). Graphical estimation of liquid composition at x_D = 0.78 and L/V = 0.78 in Example 10-28. (f). Graphical estimation of liquid composition at x_D = 0.775 and L/V = 0.78 in Example 10-28.

Table 10-30 Results of Example 10-28.

x_D	L/V	x_W	$(x_D - x_W)$		Number of ordinates
0.800	0.78	0.575	0.225	4.444	
0.795	0.78	0.445	0.350	2.857	0
0.790	0.78	0.350	0.440	2.273	1
0.785	0.78	0.255	0.530	1.886	2
0.780	0.78	0.100	0.680	1.471	3
0.775	0.78	0.080	0.695	1.439	4

Determine the values of the integral between x_i = 0.5 and various lower limits of x_w bottom composition. The area under the

curve is:

where n = number of ordinates.

x_i	x_W	[Area under the curve]
0.5	0.35	
0.5	0.255	
0.5	0.10	
0.5	0.08	

For a first attempt of $x_W = 0.35$, the area = 0.385. W is:

$D = W_i - W = 500 - 340.23$

= 159.77 kmol.

The average distillate composition, is:

For a second attempt of $x_W = 0.255$, the area = 0.5689 and W is:

$D = W_i - W = 500 - 283.1$

$= 216.9$ kmol.

The average distillate composition, is:

For a third attempt of $x_W = 0.10$, the area = 0. 84306 and W is:

$D = W_i - W = 500 - 251.2$

$= 284.8$ kmol.

The average distillate composition, is:

From the above calculations, $x_w = 0.10$ corresponds to the specification of the distillate of 0.8.

The time θ, for the batch distillation is determined from:

(10-339)

W_i = 500.0 kmol, R = 3.0, V = 80 kmol/h, e^Q = 2.323, Q = Area of $1/(x_D - x_W)$

Substituting these values in Equation (10-339) gives:

An alternative equation in determining the time (hours) for distillation is:

(10-340)

10.34.7. Batch Rectification with Variable Reflux Rate, a Fixed Number of Theoretical Plates in Column and Constant Overhead Composition

This mode of batch rectification requires the continuous adjustment of the reflux to the column to achieve a steady overhead distillate composition. Commencing with a liquid in the kettle that is rich in the more volatile component, a relatively low reflux ratio will be required to achieve the specified overhead distillate composition. This ratio must be continuously increased over time to maintain a fixed overhead composition. Ultimately, a practical maximum reflux is reached and the operation normally would be stopped to avoid distillate contamination.

Overall material balance at time θ [130, 131]:

(10-341)

(10-342)

where:

W = mols in still/bottoms at any time

W_o = mols in still/bottoms at initial charge time

Differentiating Equation (10-342) with respect to θ for varying W and x_W,

(10-343)

For constant molar overflow, the rate of distillation is given by the rate of loss of charge:

(10-344)

where D is now the amount of distillate, and not the distillation rate. Substituting Equation (10-344) into Equation (10-343) and integrating between the limits x_{Wo} and x_W, the time required for the distillation only is:

(10-345)

θ does not include the time for charging the kettle, raising the still

temperature to the bubble point, shutdown, cleaning, drainage of the product receiver, *etc.*

For fixed values of W_o, x_{Wo}, x_D, V and the number of equilibrium stages, the McCabeThiele diagram is used to determine the values of L/V for a series of values of still composition between x_{Wo} and the final value of x_W. These values are then used with Equation 10-345 to determine, by graphical or numerical integration, the time for rectification, or the time to reach any intermediate value of still composition. The required number of theoretical stages can be estimated by assuming total reflux conditions for the final value of x_W. During distillation, while rectification is proceeding, the instantaneous distillate rate will vary accordingly, and Equation 10-344 can be expressed in terms of L/V as:

(10-346)

To determine the column diameter (with trays), an approach [130] is to (1) assume θ hours; (2) solve for V, lb/h vapor up the column at selected, calculated, or assumed temperature and pressure; (3) calculate column diameter assuming a reasonable vapor velocity for the type of column internals (refer to “Mechanical Designs for Tray Performance” in this volume).

Solve for the value of θ graphically or numerically.

$1/[1 - (L/V)]\ (x_D - x_w)^2$ versus x_w, and determining the area under the curve between x_{wo} and x_w. Then substitute this value for the integral in the Equation 10-345 and solve for θ.

Figure 10-67 illustrates a distillation with variable reflux ratio; batch

operating lines with different L/V slopes all passing through x_{Di} (distillate desired overhead composition for i). Establish the McCabeThiele-like steps down each operating line until the last horizontal step or stage intersecting the equilibrium line indicates the required bottoms composition, x_{wi}.

Figure 10-67

Variable-reflux batch process solution.

Modified and used by permission, Ellerbe, R. W., Chem. Eng., May 28 (1973), p. 110.

From Figure 10-63 the separation from x_o, initial kettle volatile material to x_3 as the distillate of more volatile overhead requires three theoretical plates/stages at total reflux ratio. Using a finite reflux ratio R_4, four theoretical plates are required. The values of reflux ratio, R, can be determined from the graph with the operating line equation as,

$$y \text{ (intercept)} = x_D/(R + 1)$$

where

x_D = concentration of volatiles in the overhead distillate, mol fraction
R = reflux ratio (L/D), where L is the liquid returned as reflux
D = quantity of liquid distillate withdrawn (Figure 10-57)

The distillate percentage drawoff, P,

$$P = 100/(R + 1), \%$$

$P = 100\ (y/D_D)$, %

The values of overhead composition can be varied from x_3 of Figure 10-63 to other values as the drawoff percentage changes. As the drawoff percentage decreases, the distillate specification can be better maintained as the distillation operation continues with a fixed number of plates. For further discussion refer to References 129, 130, 131, 133.

In order to determine the time θ of distillation or the vapor load, the following procedure is as follows:

1. Draw the vapor–liquid equilibrium (VLE) diagram for the more volatile component (MVC) on the x–y plot.
2. Draw the 45°-line.
3. Locate the point corresponding to the value of x_D on the 45° line.
4. From this point draw several operating lines with varied slope (L/V) (Figure 10-67).
5. Draw the steps between the operating line and equilibrium curve, similar to that followed in the McCabeThiele method (One step is equivalent to one theoretical stage).
6. Select an operating line such that the intersection of the last horizontal step with the VLE curve gives x_W – the desired final value.
7. Plot x_W against

Determine the area under the curve between x_{Wo} and x_W. This area represents the value of the integral of the RHS of Equation 10-345.

Example 10-29 Time of Distillation at Varying Reflux Ratio and Constant Product Specification

A batch containing 500 kmols of ethanol and water mixture is to be distilled at 101.32 kPa. The initial charge contains 50 mol % ethanol, it is reduced to a residual liquor of ethanol at a concentration of 4.5 mol % using a batch distillation column with four theoretical plates including the still pot. The rate of vaporization is 80 kmol/h and the product specification is 70 mol % ethanol. Determine:

1. The time required for the distillation to achieve the desired product specification.
2. The quantity of the product obtained.

Solution

The procedure is analogous to Example 10-28 but using a variable instead of a fixed reflux ratio.

1. Select several arbitrary values of L/V ratio.
2. Draw several operating lines at L/V ratios from the 45° line for MVC (i.e. ethanol).
3. Locate the point 0.7 (x_D) on the 45° line.
4. Draw several operating lines (OL) from this point with slopes equal to the various values of L/V selected.
5. Step-off four steps along each operating line.
6. At the end of the fourth-step, determine the equilibrium value of the bottoms product.
7. Integrate:

between the limits mol % and x_W = 4.5 mol % graphically or numerically.

8. Determine the time θ from,

9. Calculate the product quantity from,

Using the vapor–liquid equilibrium data of ethanol and values of L/V, locate the point x_D = 0.7 on the 45°- line in Figure 10-67a.

Figure 10-67a VLE diagram for the mixture ethanol and water at 1 atm. : Example 10-29

Draw the operating lines at varying values of L/V from this point and step off four plates on each operating line to find the equilibrium value of the bottoms. The results are tabulated below.

L/V	x_D	x_W		
0.60	0.7	0.15	0.55	8.264
0.65	0.7	0.09	0.61	7.678
0.70	0.7	0.075	0.625	8.533
0.75	0.7	0.055	0.645	9.615
0.80	0.7	0.045	0.655	11.654
0.85	0.7	0.04	0.66	15.305

The area under the curve can be determined by:

where n = number of ordinates (starting from 0).

The area under the curve by the formula given above is:

W_o = 500 kmols, V = 80 kmol/h, x_D = 0.7, x_W = 0.5,

The time θ for the distillation to achieve the desired product specification of 70 mol % is:

= (500/80) (0.70 – 0.50) (0.939)

= 1.2 h (72 mins).

The quantity distilled is:

An Excel spreadsheet Example 10-29.xls has been developed to determine the time required for distillation and the amount distilled. Figure 10-68 shows the plot of x_W against:

with an area of 0.95. Figure 10-69 shows typical McCabeThiele plots

at varying L/V ratio and constant distillate composition at x_D = 0.7 using four theoretical stages (including the still pot).

Figure 10-68

Batch distillation of ethanol–water mixture at variable reflux and constant distillate composition.

Figure 10-69

Example 10-29: Graphical estimation of liquid composition on 4th place using variable reflux ratio.

Example 10-30 Minimum Reflux: Ratio Estimation and Associated Distillate Composition

A mixture of ethanol-water is subjected to batch distillation. Determine: (i) the minimum reflux ratio, (ii) the distillate composition; (iii) the still composition at the end of distillation. Inflection point to the VLE diagram of ethanol occurs at the point (0.7, 0.75).

Solution

From the vapor–liquid equilibrium data of ethanol, the inflection point P (0.7, 0.75) is located in Figure 10–70 and the tangent to the VLE

curve is drawn through this point and extended to meet the 45°- line at D, which corresponds to the distillate composition, x_D with the coordinates (0.82, 0.82). The tangent line DP cuts the VLE-curve at point F corresponding to the bottom composition x_W with coordinates (0.08, 0.39).

Figure 10-70

Determination of minimum reflux ratio in Example 10-30.

(i) The VLE-curve in Figure 10-70 has an inflection and the operating line for the minimum reflux is the line that passes through the point of inflection P, and is tangent to the VLE curve at this point. In Figure 10-70, the operating line for the minimum reflux ratio passes through DPF where

P = inflection point
D = Distillate composition
F = Bottom composition

The minimum reflux ratio in Figure 10-70 is:

(ii) The distillate composition is:

The still composition is determined from point F on the VLE

diagram, and the coordinates at these point correspond to the bottom composition at the end of the distillation.

(iii) The still composition is:

Example 10-31: Batch Distillation, Constant Reflux Ratio; Following the Procedure of Block [133]

Purify a mixture of ethanol and water; 11,500 lb.

Feed to kettle	Wt Per cent	lb	MW	Mols	Mole fraction
Ethanol	35	4,025	46	87.5	0.174
Water	65	7,475	18	415.27	0.826
Total	100	11,500		502.77	1.000

Overhead distillate product desired: 91.5 wt% ethanol.

Kettle bottoms residue: Not specified, as results from separation.

Vaporization rate: Assume 72 mols/h.

Average mol. weight of feed: = 11,500/502.77 = 22.87.

Solution

Overhead Product

Feed to kettle	Wt Per cent	Lb	MW	Mols	Mole fraction
Ethanol	91.5	91.5	46	1.989	0.808
Water	8.5	8.5	18	0.472	0.192
Total	100	100		2.461	1.000

Select L/V (internal reflux) = 0.75

Then: L/V = R/(R + 1) = 0.7875 = R/(R + 1), see below;

R = 3.705 (external reflux, L/D)

Because V = L + D

72 mols/h = (0.7875 V) + D

Use the ethanol–water curve similar to Figure 10-71, or refer to the data of Reference 133; the point of tangency of the line from the distillate composition of the diagonal is x_D = 0.80 and y_v = 0.80. Thus the minimum internal reflux is set by this tangent line:

Figure 10-71

x–y diagram for ethanol–water mixture, showing minimum reflux ratio.

For practical design, select L/V = (1.5) (0.525) = 0.7875.

Select L/V internal reflux lines and add to the equilibrium plot, similar to that shown for a “normal” curve of Figure 10-65.

Tabulate:

Select x_D	x_W	$(x_D - x_W)$	$1/(x_D - x_W)$
0.808			
0.770			
0.750	read, 0.085	0.665	1.5037

For a plot of x_D = 0.750, slope = 0.7875, read x_W at the equilibrium line for each theoretical tray and plot similar to Figure 10-72. Then determine the area under the curve between the selected x_w and the product x_D. Then:

(10-345a)

Figure 10-72

Graphical integration of Rayleigh, or similar, equation by Simpson's Rule for Example 10-31.

Here, ln (W_i/W) = approximately 0.210. so, W_i/W = 1.2336

Average overhead composition,

Time required for distillation only (excluding charging, heating-up, draining, cleaning, etc.):

(see right side of Equation 10-345a)

Distillation rate

Checking:

$\theta = (R + l)\,(W_i - W)/V$

$\theta = (3.705 + 1)\,(502.77 - 407.54)/72$

$\theta = 6.22$ h

where:

D = distillate rate, moles/h
L = liquid flow, moles/h
V = vapor rate, mo1es/h
= quantity of vapor, moles
W = contents of still pot, moles

x = mole fraction of substance in liquid
y = mole fraction of substance in vapor
a = relative volatility
θ = time, h

Subscripts:
D = relating to distillate
i = initial
w = relating to pot liquors

Example 10-32: Batch Distillation Calculation of Vapor Boil-up Rate for Fixed Trays (used by permission of Treybal [129]; clarification added by Ludwig [271])

Distill a small quantity each day to obtain relatively pure ortho-xylene from a mixture of ortho and para-xylene, having boiling points of 142.7°C and 138°C, respectively. The feed is 15 lb-mols (about 225 gallons) per batch, at 0.20 mol fraction para-xylene. The desired residue product is 0.020 in the kettle, while the distillate is to be 0.400 mol fraction para-xylene. A distillation column equivalent to 50 theoretical plates is to be used.

The time requirement is to complete the distillation/recovery in six hours, allowing an extra two hours for charging, emptying, and cleaning. What is the constant rate at which the distillation must be carried out?

F = 15 lb-mols/h

$x_F = 0.20$; $x_D = 0.400$; $x_w = 0.02$

The material balance:

$D = F\,(x_F - x_w)/(x_D - x_w)$

$D = 15\ (0.2 - 0.02)/(0.400 - 0.020) = 7.105$ mols

Then, $F = D + W$

$W = 15 - 7.105 = 7.895$ mols

$\theta_1 = D/G$

and, $\theta = 7.105/G$

Using Equation 10-324b,

(10-324b)

Solve for x_{Si} by trial and error.

$= 4.26$

At 138.4°C, the vapor pressures of ortho-and para-xylene are 660 and 760 mm Hg, respectively.

Because Raoult's law applies:

$\alpha = 760/660 = 1.152$

Solving the equation by trial and error shows that $x_{Si} = 0.18530$. Solving for the minimum number of plates required:

N_{min} (in column) + 1 (kettle) = 24.6

The results indicate that 25 theoretical plates are the minimum required. Assuming an efficiency of 50%, 50 total actual trays should be adequate. Choose values of k (see nomenclature) and solve for b and x_S by:

$b = (\alpha k/c) - k$

$c = 1 + (\alpha - l)k$

The tabulated results are:

k	**b**	**x_S**
0.0200	0.00297	0.01899
0.0500	0.00716	0.04842
0.0750	0.01043	0.07301
0.1000	0.01347	0.09728
0.1250	0.01631	0.12109
0.1500	0.01895	0.14470
0.1750	0.02137	0.17145
0.2000	0.02360	0.18133

Graphical integration shows the area under the curve, Figure 10-72a to be 15.311. Applying this to:

Figure 10-72a Graphical integration for boil-up rate of batch distillation for Example 10-32.

Used by permission, Treybal, R. E., Chem. Eng. Oct. 5 (1970), p. 95.

Then, $\theta_2 = 7.895\ (15.311)/G = 120.88/G$

$\theta_1 + \theta_2 = 6\ h = 7.105/G + 120.88/G$

G = 21.33 1b-mols vapor/h

This is the boilup rate, which is approximately 3.3 ft^3 vapor/sec. An approximately 1 ft. diameter column can handle this rate, where

A, B, C, E, H, J, K = constants developed in article [129]
b = y-intercept of operating line
c = constant
D = distillate, lbmoles
F = charge to batch distillation, lbmoles
G = vapor boilup rate, lbmoles/h
k = value of x at intersection of operating line and equilibrium curve
L = liquid reflux rate, lbmoles/h
N = number of ideal plates in column
N_{min} = minimum value of N
W = residue, lbmoles

x = mole fraction more volatile component in liquid
x_D = mole fraction more volatile component in final distillate
x_F = mole fraction more volatile component in feed
x_p = mole fraction more volatile component in liquid leaving column at any time
x_S = mole fraction more volatile component in kettle at any time
x_{si} = value of x_S when distillate receiver is first filled
x_w = mole fraction more volatile component in final residue
x_l = mole fraction more volatile component in distillate at any time
y = mole fraction more volatile component in vapor
y_S = mole fraction more volatile component in vapor entering column at any time
α = relative volatility
θ_1 = time for filling distillate receiver, h
θ_2 = time for refluxed distillation, hr

Example 10-33: Batch Differential Distillation of a Binary Mixture

Dimethyl ether is to be separated from a mixture with methanol. A batch operation is to be tested to evaluate whether an existing coil-in-tank can be used as a still. The pressure of the system will be about 55 psia. How many total mols will remain in the bottoms when the bottoms liquid composition contains 0.5 mol percent dimethyl ether? What is the composition of the total overhead collected?

π = system (total) pressure, psia.

= vapor pressure of component i.

	Initial Charge	At

Component	Mols	Mol Fraction	104°F Vap. Press. psia	
Dimethyl ether	61	0.427	125.0	2.27
Methanol	82	0.573	5.1	0.093
Total	143	1.000		

Initial boiling point of mixture = 104°F.

(10-347)

B_{To} = 143 mols
x_o = 0.427
x_l = 0.005
α = 125/5.1 = 24.5

B_{T1} = 143/2.15 = 67 mols remaining in bottom when dimethyl ether is 0.5 mol %.

Total vapor collected overhead = 143 − 67 = 76 mols

Mols dimethyl ether in bottoms = 0.005 (67) = 0.335

Mols dimethyl ether overhead = 76 − 0.335 = 75.665

Composition of total overhead collected:

Dimethyl ether =

Methanol = 100.0 − 99.6 = 0.4%

10.34.8. Differential Distillation-Simple Batch of a Multicomponent Mixture Without a Rectification Column

For multicomponent systems, the relation of the system can be expressed using the relative volatility:

(10-348)

where

B_i = mols of component, i, after a given time.

B_{io} = mols of component i, at start.

B_b = mols of component, b, used as reference for volatility after a given time.

B_{bo} = mols of component, b, used as reference for volatility.

Knowing the amount of components present at the beginning, the quantity remaining after the end can be calculated.

Example 10-34: Multicomponent Batch Distillation Calculation

A mixture of C_2 to C_4 hydrocarbons is to be differentially distilled at 80 psia until the concentration of propane is reduced to 10 mols per 100 mols of bottom feed material. A kettle with bottom coil is to be used, and no trays.

Material in kettle at start of distillation:

Component	Mol fraction
C_2H_6	0.1
C_3H_8	0.25
n-C_4H_{10}	0.35
i-C_4H_{10}	0.30
	1.00

Solution

Basis: 100 mols of bottoms feed.

Bubble Point of Initial Charge

Component	x_i Mol. Fract.	@ 50°F, K_i	y = Kx	Assumed 105°F, K_i	α_i @ 50°F K_i/K_p	α_i @ 105°F K_i/k_p	
C_2H_6	0.10	4.5	0.45	7.2	3.81	3.28	3.54
C_3H_8	0.25	1.18	0.295	2.2	1.0	1.0	1.0
n-C_4H_{10}	0.35	0.33	0.115	0.75	0.28	0.341	0.310
i-C_4H_{10}	0.30	0.48	0.144	1.0	0.407	0.454	0.430
Total	1.00		1.004				
			O.K.				

Note: K values at 80 psia from Natural Gasoline Supply Man's Assoc. Data Book [48]

Book [48].

Propane is reference material.

mols in bottoms

Component	B_i	Final Bottoms x_i	Vapor Press. At 105°F, psia
C_2H_6	0.39	0.00686	840
C_3H_8	10.00	0.176	200
n-C_4H_{10}	26.30	0.463	57
i-C_4H_{10}	20.20	0.355	78
	56.89	1.000	

x_i = 0.39/56.89 = 0.00686

Vapor pressure from N.K. Rector chart in Reference 48.

Second Attempt

Component	Initial x_i	α_i 50°F	Assume 95°F, K_i	α_i @ 95°F K_i/K_p	
C_2H_6	0.10	3.81	6.7	3.35	3.58
C_3H_8	0.25	1.0	2.0	1.0	1.0
n-C_4H_{10}	0.35	0.28	0.67	0.335	0.307
i-C_4H_{10}	0.30	0.407	0.92	0.46	0.432

Component	Final B_i	Vapor Press.@ 95°F, psia	Final x_i	
C_2H_6	0.378	750	0.0066	4
C_3H_8	10.0	174	0.175	3
n-C_4H_{10}	26.4	47	0.464	2
i-C_4H_{10}	20.2	67	0.354	2
Total	56.978			8

Therefore the final temperature should be close to 95°F, because 80.7 psia is close to the operating pressure of 80 psia.

Total mols of bottoms remaining at end: 57 mols liquid

Total mols vaporized = 100 − 57 = 43

Liquid composition mol fraction is given in column “Final x_i,” and corresponds to the actual mols B_i, noting that there are the required 10 mols of propane in the bottoms under these conditions.

10.34.9. Batch Distillation with a Fractionation Column-Constant Overhead Product Composition. Multicomponent or Binary Mixture

The method of Bogart [4] is useful in this case. The basic relation is:

(10-349)

Application may be (1) to determine a column diameter and number of plates or (2) with an existing column and an assumed operating reflux, determine the time to separate a desired cut or product, where:

θ = time from start when given L/V will produce constant overhead composition, x_{iD}

B_{To} = mols total batch charge to still

V = total mols per hour vapor overhead

x_{iD} = mol fraction light key component in overhead product

x_{iB} = mol fraction light key component in original charge

Procedure for situation (2) above, using an existing column:

1. Calculate the minimum number of plates and minimum reflux ratio:
   a. For multicomponent mixture, select key components, light and heavy.
   b. Calculate relative volatility, α_i, referenced to heavy key component, at top and bottom temperatures, and determine geometric average α.
   c. Calculate minimum theoretical plates at total reflux by Fenske's equation (10-182).
   d. Use Gilliland correlation to determine actual reflux ratio, using an estimated number of actual plates, and a minimum reflux ratio from:

e. Calculate:

Internal (L/V) =

2. Set up table: Keep x_{lD} values constant.

*Assumed "x_1" values	(L/V)	(1 − L/V)	(x_{1D} − x_1*)	A	B
x (bottoms)	•	•	•	•	•
x_2	•	•	•	•	•
x_3		•	•	•	•
:	•	•	•	•	•
x (feed)	•	•	•	•	•
A = (1 − L/V) (x_{1D} − x_1*)2					
*Assume "x_l" values of bottoms compositions of light key for approximate equal increments from final bottoms to initial feed charge. Calculate L/V values corresponding to the assumed "x_l" values by inserting the various "x_l" values in the Fenske equation for minimum reflux ratio of l-(d). The "x_l" values replace the x_{lB} of this relation as the various assumptions are calculated. The actual (L/D) are calculated as in l-(d) keeping the number of trays constant. Complete the table values.					

3. Plot:

The total area under the curve ΣA, may be obtained in several ways; the rectangular or trapezoidal rules are generally satisfactory. The area is that between the original feed and the final bottoms composition for the particular component.

4. Time required for a batch:

(10-350)

V is an assumed or known value, based on reboiler capacity.

5. Plot of reflux ratio quantity versus time.

From the L/D values of l-(e), knowing the L/V, using V assumed as constant, calculate the necessity reflux fluid, L. Figure 10-73 indicates a plot of time to produce a constant product composition and the necessary external reflux returned to the tower.

Figure 10-73

Batch distillation with trays: constant overhead product.

The batch distillation calculation of a binary is simpler, as L/V values can be assumed, and because there is only enrichment of the overhead involved only one operating line is used per operating condition. The number of theoretical trays can be stepped-off and x_{iB} values read to correspond. The plots involved are the same as previously described. Table 10-31 provides rules of thumb for batch distillation

Table 10-31 Heuristics on Batch Distillation.

• Too low a reflux ratio cannot produce the required product specification no matter how many trays are installed. Conversely, even infinite reflux will not be sufficient if an inadequate number of equilibrium stages has been provided.

• For optimum separation efficiency, reflux holdup should be minimized by eliminating surge drums and using flow splitters that retain little or no liquid.
• Too little or too much holdup in the column is detrimental to separation efficiency. A reasonable amount provides a flywheel effect that dampens the effects of equilibrium-condition fluctuations; too much, especially at higher reflux ratios, makes it difficult to achieve good purity levels. A hold-up equivalent to 10–15% of the initial batch charge is recommended.
• Since the column consists solely of a rectifying section, there is a limit to how much trays can be profitably installed. The system will "pinch" regardless of stages once the low boiler concentration in the reboiler approaches the intersection of the operating line with the equilibrium curve.
• Once a workable column has been installed, capacity to produce at a given rate and product specification is only minimally affected by changes in reflux ratio or length of a cut.
• As the more-volatile component is being removed from the reboiler, separation becomes progressively more difficult.
• It is impossible to recover in a single operation, at high purity, a low boiling component that represents only a small fraction of the initial charge.
Source: Chemical Engineering magazine, SRI Consulting: A division of Access intelligence, LLC.

Rigorous Methods

The rigorous methods for calculation of batch distillation operations are rather more complex than for continuous steady-state distillation,

because of their transient operation. The basic assumption of steady state operation in the continuous column model is not applicable in these instances. The only operation for steady state in batch distillation is at total reflux, which is used as the initial condition for the dynamic solution of the column.

The rigorous solution involves coupling both the heat and material balance and phase equilibrium equations. Hold-up terms must also be considered. Distefano [281] assumed a constant volume of liquid holdup, negligible vapor holdup, and negligible fluid dynamic lag. He describes a computational procedure to solve the equations one at a time. Boston *et al.* [282] describe a more flexible and efficient method that can handle a variety of specifications and column configurations, including side draws, side heaters and multiple feeds. Their method uses advanced numerical procedures to handle nonlinearity and stability problems. Khoury [283] has outlined Distefano's method in order to illustrate some of the potential numerical computational difficulties that are encountered. Further details of these methods are provided by Distefano and Boston *et al.*

10.35. Steam Distillation

Live steam in steam distillation, is fluids being distilled, either batch or continuous.

Ellerbe [127, 128] provides an excellent summary of the basics of steam distillation. The theory of direct steam distillation evolves around the partial pressures of the immiscible organics/petroleum/petroleum components and the presence of direct open steam in the system. The system may consist of the organic immiscible plus steam (vapor and/or liquid). Each liquid exerts its own vapor pressure independent of the other. The total pressure of the system is hence the sum of the individual vapor pressures of the two liquids (assuming the liquids are immiscible). An important use of this

approach is to separate a volatile organic from non-organic impurities.

At constant temperature, the partial pressure for each component, and the composition of the vapor phase, are independent of the mols of liquid water or organic compound present. For example, for a system held at 800 mm Hg, the mixture could boil at, 250°F, and both liquids present would boil over together. Should one component evaporate (boil away) before the other, the system' vapor pressure would then fall to the temperature corresponding to the boiling point of the residual material.

For a system such as discussed here, the Gibb's Phase Rule [59] applies and establishes the "degrees of freedom" for control and operation of the system at equilibrium. The number of independent variables that can be defined for a system is expressed as:

(10-351)

where

ϕ = number of phases present
F = degrees of freedom
C = number of components present

For example, for steam (saturated vapor, no liquid) distillation with one organic compound (liquid), there are two phases, two components, and hence no degrees of freedom. The degrees of freedom that could be set for the system could be: (1) temperature and (2) pressure; or (1) temperature and/or (2) concentration of the system components; or either (1) pressure and (2) concentration. In steam distillation, steam may be produced from the water present, so both liquid and vapor phase water (steam) are present. For such a case, the degrees of freedom are $F = 2 + 2 - 3 = 1$.

The basic calculations involve the partial pressures of the components as discussed earlier.

For batch steam distillation: stripping [127, 128]

(10-352)

π = total system pressure (also see Equation 10-3)

$y_{im} = p_{im}/\pi$

The steam required for the distillation is:

(10-353)

(also see Equation 10-2)

The steam required per mol of immiscible liquid vaporized is:

(10-354)

(10-355)

When the sum of the partial pressures of the steam and the material

distilled reach the system pressure, boiling begins and both components are vaporized in the mol ratio of their partial pressures. Upon condensation of the overhead mixture, the condensate receiver will contain two layers separated by gravity.

The weight ratio of steam to the immiscible liquid in the vapor is:

(10-356)

Any nonvolatile material in the mixture will remain in the still bottoms. The Hausbrand's vapor-pressure diagram [127, 128] in Figure 10-74 is a useful approach for the steam distillation calculation. This diagram was prepared for six organic compounds and the corresponding water vapor pressure as $(\pi - P_s)$ for three system pressures of 760, 300, and 70 mm Hg versus temperature, where

M = molecular weight of material
p = partial pressure, mm Hg
W = weight of material in vapor
N = number of mols
N_o = number of mols of nonvolatile material present
y = mol fraction of material in vapor
π = system pressure, mm Hg
p_{im} = pure component vapor pressure of the immiscible liquid being distilled

Figure 10-74

Hausbrand vapor-pressure diagram for various liquids and at three system steam pressures. A similar diagram can be constructed for other organic/hydrocarbon systems.

Used by permission, Ellerbe, R. W., Chem. Eng. Mar. 4 (1974), p. 108.

Subscripts:

im = immiscible liquid
s = steam
1 = initial
2 = remaining

The water curve intersects the particular organic compound, and at that point the temperature is that at which the steam distillation can take place, because the partial pressures are additive at this point. For example, for 300 mm Hg total pressure system, reading the intersection point for benzene, and steam at 46°C, gives 220 mm Hg for benzene and (300 − 220) or 80 mm Hg. Then the mol ratio of benzene to water vapor is therefore 220/80 = 2.75; or 2.75 parts of benzene to 1 part of water.

When the composition of the compounds in the still or bottoms changes significantly as the batch distillation progresses, an unsteady state condition will exist as for differential distillation.

When nonvolatile material in the bottoms is significant, and no liquid water exists there – that is, p_s is below the saturation of the steam pressure at the still temperature – then the Raoult's Law steam efficiency is [127]: Values of E are found to range from 60% to 70% for many organics, but values of 90% to 95% are reported [127] for good sparger design for steam injection, and molecular weight of organics <100, and 50% for many lubricating oils.

(10-357)

where

E = vaporization efficiency of steam distillation

Note that for this discussion now, P_{im} just above and in equations to follow, refers to the pure component vapor pressure of the immiscible liquid being distilled [127].

When steam is added to the still [127]:

(10-358)

and, for a constant distillation temperature, p_{im} is constant. Then for constant pressure:

(10-359)

As the volatile material is reduced over the batch distillation process, the steam pressure rises during the progress of the operation, due to the decrease of p_{im}. When the volatile material is stripped down to a low residual concentration, p_s approaches the total system pressure, π. When the steam saturation pressure with temperature is $>\pi$, no steam condensation will occur during the operation.

When the nonvolatile concentration is low to insignificant in the still feed, then N_o is small relative to N_{im}. Then p_{im} is considered constant [127]:

(10-360)

(10-361)

Then, the distillate:

(10-362)

Operation of the open still with only a steam injection sparger to bring steam below the liquid level in the still may not be efficient as operating the unit with internal trays to act as a stripping column. This should be examined for each situation, as the installation of trays can be expensive, particularly if they do not aid significantly in achieving the desired separation. Generally, when no liquid water is present, the best operation of an open still (with condenser) may be at the highest working temperature achievable without adverse effects on the fluids (i.e., no polymer formation, no breakdown, etc.). Often, direct steam injection can be reduced for any operation by careful heating of the still by an internal reboiler coil, or possibly a steam jacket. The heat sensitivity of the compounds involved must be recognized.

10.35.1. Steam Distillation-Continuous Flash Operation, Multicomponent or Binary Mixture

This system requires direct steam injection into the still with the liquid; all the steam leaves overhead with the boiled-up vapors (no internal condensation) in a steady-state operation, and system at its dew point. Steam is assumed to be immiscible with the organics. Steam distillation is usually applied in systems containing high boiling organics, or heat sensitive materials which otherwise require

separation under vacuum conditions.

(10-363)

b is more volatile reference component

$i \neq s$ = components, i, are not to include steam, s

M_s = total mols steam required

B_{To} = total mols hydrocarbons at start (not including the steam)

B_{Tl} = mols liquid in bottoms of still at time, T

B_{io} = mols of component, i, at start

α_i = relative volatility of more volatile to each of other components

P_b = vapor pressure of reference more volatile component, b

π = total system pressure, absolute

B_b = mols of component, b, used as reference for volatility, after a given time of distillation

B_o = mols of component, b, used as reference for volatility, at start of distillation

Example 10-35: Multicomponent Steam Flash Calculations

A mixture of bottoms material of composition B_{io} below has accumulated in the run-down tank. It is necessary to separate the volatile organic heavies from the tarry polymerized residue (heavy liquid). Steam is to be injected into the insulated tank containing heating coils. The system is to operate at 200 mm Hg absolute pressure and 250°F with no condensation of the steam. The organic volatile heavies contain:

Component	**Vapor pressure @ 250°F, mm Hg**	**$\alpha = p_i/P_A$**	**Mols B_{io}**	**B_{io}/α_i**
A		1.0	45	45
B	20	0.57	40	70
C	6	0.171	26	152

Total			111	267

= 13.75 – 1 = 12.75

M_s = 1,415 total mols steam required for 111 mols mixture

Mols steam/mol of mixture organic volatiles = 1,415/111 = 12.8

10.35.2. Steam Distillation-Continuous Differential, Multicomponent or Binary Mixture

The results of the differential distillation are the same as the flash distillation, although the mechanism is somewhat different. This is a batch type operation distilling differentially. All sensible and latent heats are supplied separately from the steam or by superheat in the steam. Steam acts as an inert in the vapor phase, and the quantity will vary as the distillation proceeds, whilst temperature and pressure are maintained.

(10-364)

If all the volatile materials are distilled: and $B_{T1} = 0$

This relation is handled in a similar manner to that in flash steam separation.

If all of the material is not to be removed as overhead vapors from the still, leave a percentage of a particular compound in the bottoms, then select the particular compound as the reference material “b” for α determinations.

B_b = (Fraction retained) (B_{bo}) and = (Fraction retained)

Substitute and solve for B_{T1}.

(10-365)

Knowing B_{Tl}, the relation for M_s can be solved to determine mols of steam to reduce initial material to percentage of a compound in the remaining bottoms. If steam condenses, the requirement for steam increases by this amount.

10.35.3. Steam Distillation-Continuous Flash, Two Liquid Phases, Multicomponent and Binary Mixture

Because water will be present in this system, and is assumed immiscible with the other components, it will exert its own vapor pressure. This situation is similar to that with many systems where the liquid to be flashed enters below its dew point, and hence requires the use of steam for heating (sensible + latent) as well as for the partial pressure effect.

Mols steam in vapor phase only:

M_s (vapor) = P_s (at assumed flash temperature)

where

P_s = vapor pressure of steam

P_i = vapor pressure of each component at the flash temperature

Mols steam to heat is sum of sensible plus latent

Total mols steam is sum of M_s (vapor) plus heating steam

System total pressure:

(10-366)

B_{To} = mols (total) volatile material at start

10.35.4. Open Live Steam Distillation-With Fractionation Trays, Binary Mixture

Open or direct injection of steam into a distillation system at the bottom of the column or still may be used to heat the mixture, as well as to reduce the effective partial pressure of the other materials. In general, if steam is used to replace a reboiler, one tray is added to replace the reboiler stage, and from 0.3 – 1.0 more trays may be needed to offset the dilution of the system with water in the lower portion. If steam is acceptable, it replaces the cost of a reboiler and any cleaning associated with this equipment. For most columns, extra trays can be purchased to offset this cost. When one of the components of the binary is water, and steam is used, the equation used for the operating stripping line (there is no rectifying section) is:

For component not including water:

$$V_s\, y_{i\,(m)} = L_s\, x_{i\,(m+1)} - B\, x_{iB}$$

Slope of operating line $(L/V)_m = B/S$

Operating line intersects the x-axis at x_{iB}

The step-off of trays starts at x_{iB} on the x-axis, y = 0.

Open steam is used for stripping of dissolved or absorbed gases from an absorption oil, with all of the steam going overhead, and the stripped oil leaving at the bottom. The absorption coefficient of the oil for the component must be known to be able to construct the equilibrium curve. The operating curve is constructed from several point material balances around the desired component, omitting the oil provided; its volatility is very low. The trays can be stepped-off from a plot of y vs. x as in other binary distillations, but using only the stripping section.

Example 10-36: Continuous Steam Flash Separation Process: Separation of NonVolatile Component from Organic Components

It is desired to separate a nonvolatile material from an equimolal mixture of benzene, toluene, and xylene at 80°C. Vapor pressure data for these compounds are available from several physical property sources. The following approximate values for the specific heats and latent heats of vaporization may be used:

Benzene: C_p = 0.419 cal/gm-°C

ΔH_v = 97.47 cal/gm

Toluene: C_p = 0.44 cal/gm-°C

ΔH_v = 86.53 cal/gm

Xylene: C_p = 0.40 cal/gm-°C

ΔH_v = 82.87 cal/gm

If the mixture is separated by a continuous flash process, the

components are considered insoluble in water (check references) and the feed enters at the flash chamber at 20°C, calculate the mols of steam condensed, the total mols steam required per 100 mols of feed, and the total pressure. Use steam at 212°F and atmospheric pressure.

In summary, this involves a continuous flash process:

- All the feed is to be flashed.
- Steam does the heating.
- Some steam condenses.
- Water is immiscible with the materials.

Solution

Feed:

Benzene:	33.33 + mols
Toluene:	33.33 + mols
Xylene:	33.33 + mols
Total	100.00 mols feed

Because water will be present in liquid phase, it will only exert its vapor pressure. Temperature of flash = 80°C.

where

N_2 = mols steam in vapor only

P_s = vapor pressure of steam

= mols of each component at start

P_i = vapor pressure of each component at temperature

Component	Mols at start	P_i at 80°C, mm Hg	
Benzene	33.33	760	0.043
Toluene	33.33	280	0.1190
Xylene	33.33	120	0.277
			0.4397

P_s at 80°C (176°F) = 6.868 lb/sq in. abs (from steam tables)

= 354 (0.4397) = 155.7 mols steam in vapor per 100 mols of feed (volatile) material

Steam required to heat feed to 80°C:

Benzene: Sensible heat	Btu
(78) (33.33) [(0.419 cal/gm-°C) (1.8)](80°−20°)	= 117,800
Latent heat	
(78) (33.33) (97.46×1.8)	= 454,000
Toluene: Sensible heat	
(92) (33.33) [0.44×1.8] (80°−20°)	= 145,600
Latent heat	

(92) (33.33) (86.53×1.8)	= 477,000
Xylene: Sensible heat	
(106) (33.33) [0.40×1.8] (80°−20°)	= 153,000
Latent heat	
(106) (33.33) (82.87×1.8)	= 525,000
Total heat load:	= 1,872,400

lbs steam required for heat load =

= 1,932 lbs steam/100 mols volatile

Mols steam required for heat load = mols steam/100 mols volatile material

Total mols steam/100 mols volatile feed = 155.7 + 107.2 = 262.9

Total pressure of system:

L_T = Total mols volatile material at start = 100

Example 10-37: Open Steam Stripping of Heavy Absorber Rich Oil of Light Hydrocarbon Content (used by permission following the method of R. W. Ellerbe, Chemical Engineering [127])

A gas processing plant selectively extracts ethylene and ethane from an incoming natural gas mixture stream. These two light hydrocarbons are absorbed in heavy gasoline type absorber “oil,” and then stripped

with open steam in an open tower. The system data are (see Figure 10-75):

Rich oil rate to tower:	8,500 mol/h
Overhead product of ethylene and ethane:	775 mol/h
Overhead product from accumulator:	55% vapor and 45% liquid
Accumulator conditions:	48 psia and 135°F
Reflux hydrocarbon in top vapor @ 175°F:	850 mol/h
Steam (superheated) enters bottoms below tray:	14,000 lb/h
Water partial pressure in the mixed vapor at bottoms:	20 psi
Hydrocarbons mix partial pressure:	50 psi − 20 psi = 30 psi

Figure 10-75

Open steam stripping light hydrocarbons from rich oil.

Modified for Example 10-37 and used by permission, Ellerbe, R. W., Chem. Eng. Mar. 4 (1974), p. 108.

Neglect pressure drops through the system.

Determine: How much water is removed from the overhead accumulator and the intermediate dehydrator or water removal tray? No water is removed from the bottoms due to the use of superheated steam.

Solution

From steam tables (saturated) at:

Top of tower, 175°F, vapor pressure water, psia	= 6.8
Mol fraction water vapor at top of tower: 6.8/48 psia	= 0.1416
Mol fraction hydrocarbon at top of tower: 1 − 0.1416	= 0.8584
Total mols mix HC vapor and water vapor at tower overhead:	
Mols of water vapor in tower overhead: 1,893 − (775 + 850)	= 268
Accumulator @ 48 psia & 135°F, water vapor pressure:	= 2.6 psia
Mol fraction water in accumulator vapor: = 2.6/48	= 0.0541
Mol fraction HC in accumulator vapor: = 1 − 0.0541	= 0.9459
Total mols vapor leaving accumulator: =	= 450.6
Mols water vapor leaving accumulator: = 450.6 − (775) (0.55)	= 24.35
Mols liquid water withdrawn from accumulator: = 268 − 24.35	= 243.65
Mols liquid water collected on dehydrator tray and removed at that point up tower above where reflux returns below this tray: (water vapor in tower overhead)	= 777. 7 − 268 = 509.7 mols/h
Mols steam entering tower: = 14,000/18	= 777.7 mols/hr

10.36. Distillation with Heat Balance of Component Mixture

This type of evaluation of a distillation system involves a material and heat balance around each tray. This is extremely tedious by conventional means, but is now handled with computers using simulation software (e.g. Hysys, Chemcad, Prosim, UniSim, Aspen). However, the volume of calculations is large and requires a relatively long time. Only those special systems that defy a reasonable and apparently economical solution by other approaches are even considered for this type of approach. The detailed method involves trial and error assumptions of both material and heat balance.

10.36.1. Unequal Molal Overflow

This is another way of expressing that the heat load from tray to tray is varying in the column to such an extent as to make the usual simplifying assumption of equal molal overflow invalid. The relations to follow do not include heats of mixing. In general they apply to most hydrocarbon systems.

1. Equation of operating line in rectifying section, light component [59]:

(10-367)

2. Equation of operating line in stripping section, light component:

(10-368)

where

$M_B = h_B - Q_B/B$

$M_D = Q_c/D + h_D$

H_n = total molal enthalpy of vapor at conditions of plate n, $H_n = \Sigma H_{ni} (y_{ni})$

h_n = total molal enthalpy of liquid at conditions of plate n, $h_n = \Sigma h_{ni} (x_{ni})$

s = lb (or mols) steam per lb (or mol) bottoms

H_m = total molal enthalpy of vapor at plate m (below feed)

N = mols residue or bottoms per unit time

Q_B = heat added in still or bottoms

10.36.2. Ponchon-Savarit Method-Binary Mixtures

For a binary fractionating tower, illustrated in Figure 10-76, the conditions of the feed, F, x_F, H_F are known and the composition of the distillate product, x_D and the composition of the bottoms product, x_B are specified. If the column is operated at a constant reflux ratio, the number of equilibrium stages required to achieve a desired separation can be estimated. The simplest method of achieving this task is based upon an assumption of constant molal overflow (CMO). This only occurs if:

1. The molal heats of vaporization of the components are identical.
2. Sensible heat distributions are negligible.
3. Heats of mixing are negligible.

Figure 10-76

Fractionating tower with reboiler and condenser units.

With these criteria, the internal energy balances around each plate in the tower are straightforward, as the molal flow rate of liquid and vapor are constant throughout each section. The McCabeThiele diagram can be employed to calculate the number of equilibrium stages. However, the above criteria are not met for numerous systems. The use of McCabeThiele analysis in these circumstances may lead to under specification of distillation equipment.

Where constant molal overflow is not applicable, enthalpy-composition diagrams can be combined with vapor-liquid equilibrium data and the graphical Ponchon-Savarit method employed. A Ponchon-Savarit analysis does not assume the liquid and vapor flow rates are constant in each section of the tower. The only assumption of this approach is that there are no significant heat losses through the column walls.

This graphical method allows solution of many distillation problems which would require considerable work if attempted by rigorous methods. Robinson and Gilliland have published technical and descriptive details substantiating this method [8, 59]. Figure 10-77 presents a summary of the use of this method and appropriate interpretations. Scheiman [104] uses the Ponchon-Savarit diagrams to determine minimum reflux by heat balances. Campagne [216, 217] suggests a detailed technique for using the Ponchon-Savarit method in a computer simulation, which leads to designs not previously possible.

Figure 10-77A–E

Performance analysis of unequal molal overflow for binary system using Ponchon-Savarit method.

The basic method allows the nonideal heat effects of the system to be considered as they affect the plate-to-plate performance. The systems, as represented in the diagrams, are usually at constant pressure, but this is not necessarily the case in reality. The equilibrium tie lines connect points fixed by the x-y values to corresponding saturated liquid and saturated vapor conditions at a constant temperature, such as “a” °F or “b” °F. The mol fractions are obtained from the usual x-y diagram for the system, and the enthalpy values are relative to a fixed datum point for the available heat data of the particular components. For such systems as ammonia-water, methanol-water and ethanol-water the data are readily available (see Figure 10-78aFigure 10-79 and Figure 10-80). The saturated liquid line represents the enthalpies of liquid mixtures at the various compositions, all at a constant pressure. This is the bubble point curve. The dew point curve is produced by plotting the enthalpies of the various vapor mixtures at the saturation temperature at a constant pressure. Figure 10-78b shows the corresponding vapor-liquid equilibrium diagram for ammonia-water system at 1.10325 kPa, and Figure 10-81 illustrates the combined enthalpy-concentration diagrams for ethanol-water system at 1.01325 kPa.

Figure 10-78a Enthaply vs composition of ammonia–water system.

Figure 10-78b VLE data of ammonia-water system at 1atm.

Figure 10-79

Enthalpy composition diagram for methanol–water mixture at a pressure of 101.235 kPa.

Figure 10-80

Enthalpy composition diagram for ethanol–water mixture at a pressure of 101.235 kPa.

Figure 10-81

Combined enthalpy–concentration diagrams for ethanol–water system at 1.0

bar.

An effort has been made to present the basic understanding of the method as it applies to systems involving unequal molal overflow, open steam distillation and single flash vaporization in Figure 10-77A–E and Figure 10-82A–C.

Figure 10-82A–C

Graphical solution of unequal molal overflow, binary systems.

To obtain high or even necessary levels of accuracy for some design conditions, the end portions of the graphical representation may require enlargement from the usual size for graphical plotting. In most cases, a size of 11 × 17 inches is suggested. In general, the nonequimolal overflow (NEMO) model of the column is more rigorous, and should be used in those columns where there are:

1. Large temperature differences from the top of the column to the bottom (i.e. where sensible heat effects are important).
2. Large heats of mixing.
3. Significant differences in the molal heats of vaporization of the components.

The Equations

Consider the energy balance around the column, assuming no heat loss through the column wall:

(10-369)

where and are the duties of the reboiler and condenser respectively, Btu/h. Since the column reflux ratio R is set, R is defined by:

(10-370)

or

(10-371)

with L = the molal flow rate of reflux returned to the column, the condenser duty is:

(10-372)

Substituting Equation (10-371) into Equation (10-372) gives:

(10-373)

where

λ is the latent heat of vaporization of the distillate product.

The distillate product flow rate, D and the bottoms product flow rate B can be determined from an overall mass balance and a mass balance on the more volatile component as follows:

(10-374)

(10-375)

Since the feed stream and the composition of distillate and bottoms product have been specified, the two unknown quantities, D and B in Equations 10-374 and 10-375 can be found.

Combining Equations 10-374 and 10-375 gives:

(10-376)

Rearranging Equation 10-376 gives:

(10-377)

From the heat balance of Equation 10-369,

(10-369)

Rearranging Equation 10-369 gives

(10-378)

Equation 10-377 in the ratio of B/D gives

(10-379)

Substituting Equation 10-374 into Equation 10-378 gives

(10-380)

or

(10-381)

Substituting Equation 10-379 into Equation 10-381 gives:

(10-382)

(10-383)

Rearranging Equation 10-383 gives

(10-384)

and

(10-385)

Equation 10-385 indicates that the points and must all lie on a single straight line on an enthalpy-composition diagram. Since we know the condenser duty, , the point can be positioned on the enthalpy-composition diagram as point d in Figure 10-83 a-c. Place the feed point f at h_F, x_F and extend a line through points d and f to a vertical line placed at $x = x_B$ to point b. This point must have the coordinates .

Figure 10-83A–C

Ponchon analysis of a fractionating column.

Note: the feed is taken as saturated liquid.

Proceed to calculate the number of equilibrium stages in the rectifying section as follows: First, draw a vertical line at $x = x_D$ extending from

point d to the saturated line at point l_0. Both the distillate product and reflux are saturated liquid with composition x_D and hence must be represented by point l_0 (Figure 10-83 b).

The mass balance, component balance and the energy balance around the condenser are:

(10-386)

(10-387)

(10-388)

Around the top plate of the column and the condenser, an energy balance yields:

(10-389)

where

h_v = enthalpy of saturated vapor

h_L = enthalpy of saturated liquid

The total and component mass balances are:

(10-390)

(10-391)

Substituting V_2 in Equation 10-390 into Equations 10-389 and 10-391 respectively gives:

(10-392)

(10-393)

Rearranging Equation 10-393 gives:

(10-394)

Rearranging Equation 10-392 gives:

(10-395)

or

(10-396)

(10-397)

Substituting Equation 10-397 into the energy balance Equation 10-394 gives

(10-398)

or

(10-399)

(10-400)

The familiar form of Equation 10-400 indicates that points d and l_1 in Figure 10-83 b must be on the same straight line as the point y_2, h_{v2}. Since the location of points d and l_1, is known a line can be drawn connecting these points. Stream V_2 is saturated vapor, and the intersection of line d–l_1 with the saturated vapor line at point V_2 must represent the enthalpy and the composition of stream V_2 on Figure 10-83 b. The process is repeated to calculate the enthalpy and composition of streams L_2 and V_3. A tie line is drawn from V_2 to the saturated liquid line. Point l_2 thus determined, specifies stream L_2. From point l_2 a line is constructed to point d. By reasoning analogous to that following Equation 10-400, stream V_3 is represented by the intersection of line l_2d with the saturated vapor line at point V_3. The process is continued for each plate in the rectifying section of the column and the number of plates determined by the number of tie

lines which were constructed.

When a tie line crosses line *bfd* (Figure 10-83), the feed point is determined. A new set of material and energy balances is then used to find the number of plates in the stripping section. This corresponds to using different operating lines in the stripping and rectifying sections of a column in a McCabeThiele analysis. It is not essential that the feed plate is chosen whose tie line crosses line *bfd* as the feed plate. However, with this criterion for locating the feed plate, the maximum separation per plate is achieved.

Stripping Section (see Figure 10-76):

The number of plates in the stripping section of the column (i.e. plates below the feed point) is determined in the same manner as in the rectifying section and the procedure is as follows.

An energy balance is written around the second plate in the stripping section and the reboiler:

(10-401)

The mass and component balances are:

(10-402)

(10-403)

Combining Equations 10-402 and 10-403 gives:

(10-404)

Rearranging Equation 10-404 gives:

(10-405)

Therefore

(10-406)

From the heat balance:

(10-407)

or

(10-408)

Substituting Equation 10-406 into Equation 10-408 gives:

(10-409)

Rearranging Equation 10-409 gives:

(10-410)

or

(10-411)

Rearranging Equation 10-411 for the stripping section of the column finally gives:

(10-412)

In the stripping section, from Equation 10-412, points b and l_4 must lie on a straight line with points H_{V5} and y_5. Since the location of points b and l_4 is known, a line is extended through these points to the saturated vapor line to determine points V_5 and the enthalpy and composition of a stream V_5 as shown in Figure 10-83 c.

Streams V_5 and L_5 are in equilibrium, and a tie line from point v_5 to the saturated liquid line set point l_5 and the conditions of stream L_5. This process is continued by drawing a line through points b and l_5 to find point v_6 on the saturated vapor line. By repeating this process until a tie line intersects the saturated liquid line at or below x_B, we complete the Ponchon-Savarit analysis. The number of equilibrium stages in the fractionating column is determined by counting the number of tie lines that were constructed. In Figure 10-83 c, a total of 7 equilibrium stages, three plates in the rectifying section, one feed plate and three plates in the stripping section corresponds to the column reboiler. Points d and b in Figures 10-83 a-c are referred to as difference points and lines drawn to d and b are referred to as

difference lines.

Minimum Reflux

The condition of minimum reflux is when any equilibrium tie line coincides with any operating line, as shown in Figure 10-84. Usually this coincidence takes place on the operating line through the feed condition, and the location of d for the condition of minimum reflux is found by extending the equilibrium tie line through the feed condition to intersect the composition through x_D or y_D.

Figure 10-84

Minimum reflux ratio on an enthalpy–concentration diagram.

Total Reflux

The condition of total reflux is shown in Figure 10-85. In this instance, the operating lines are parallel and vertical, converging at infinity. This results from the condition of V = L since D = 0, B = 0, , and . Thus, the enthalpy coordinates are and . The minimum number of plates is encountered at the condition of total reflux.

Figure 10-85

Total reflux construction on an enthalpy–concentration diagram.

Side streams and multiple feeds

The Ponchon-Savarit method is very versatile and can be applied with columns having multiple feeds or side stream withdrawals. There are situations where feeds of different composition are involved, and it is possible to introduce each at its proper location in the column, and thereby reduce the total number of plates by one or two compared when the feeds are premixed. Alternatively, process requirements may require two or more distillate products of different composition, and these may be removed from the column as distillate and side streams. The Ponchon-Savarit method can be employed in these cases by writing the difference-point equations for each section of the column between streams added or removed. The number of difference points will always be equal to the number of streams N (including the distillate and bottoms) minus 1 as shown in Figure 10-86 and Figure 10-87.

Figure 10-86

Ponchon analysis of a column with Multi feed and with side stream.

Figure 10-87

Ponchon analysis of a column with two feed streams

Columns with side streams may be handled in much the same way as a single feed stream. Here, three difference points are needed to represent the three sections of the column in Figure 10-86. Point d_1 and f′ are located as shown, with f′ representing a hypothetical stream in which a side stream of saturated liquid has been physically subtracted from the saturated liquid feed. Using the lever-arm principle, point f′ must lie on a line extended through points s and f such that . Similarly, in the case of multiple feeds, the difference point b is found by the intersection of the line d_1 f′ and a vertical line at x_B. The difference point for the intermediate section d_2, is determined by the intersection of lines sb and fd_1. The number of equilibrium stages may be stepped off as usual.

The advantage of the H-x-y chart lies in the fact that the heat quantities required for the distillation are clearly indicated. Therefore, the higher the reflux ratio, the more heat must be removed per mole of product. This immediately shows that both q_c and Q_B are increased. The Ponchon-Savarit method is applicable to a wide range of binary flash and distillation problems. The method is more rigorous and may be used for a larger class of binary systems than the McCabeThiele method. However, the final choice between the Ponchon-Savarit and McCabeThiele methods must be based on the desired accuracy of results and the availability of data.

Example 10-38: Ponchon-Savarit Unequal Molal Overflow in Ammonia–Water System

An ammonia–water recirculating solution of 62 wt % is to be stripped of the ammonia, for recovery by condensation at 260 psia with river

water cooling. The overhead ammonia product is to be at least 99.5 wt % and the bottoms should approach 0.05 wt % ammonia. The feed enters as a liquid at its boiling point, with an enthalpy of 42 Btu/lb.

Enthalpy Diagram

Prepared by reading the h and H values from the Jennings and Shannon Aqua-Ammonia Tables [35] at 260 psia and various wt %'s of ammonia in the liquid. The tie lines connect the vapor compositions with the equilibrium liquid values, Figure 10-88.

Figure 10-88

Ponchon type diagram for ammonia–water distillation.

Vapor–Liquid Equilibrium Diagram

Prepared from corresponding x and y values in Reference 35 at 260 psia, Figure 10-89.

Figure 10-89

McCabeThiele diagram for ammonia water system.

Number of Trays

x_F = 0.62 weight fraction ammonia

x_D = 0.995 weight fraction ammonia

x_B = 0.0005 weight fraction ammonia

1. Minimum Reflux

(10-413)

From enthalpy-composition diagram:

H_D = 590 Btu/lb

h_D = 92 Btu/lb (assuming no subcooling)

$(M_D)_{min}$ = 596 Btu/lb

$(M_D)_{min}$ is determined by reading the equilibrium y value corresponding to the feed composition 0.62 from the x-y diagram, noting it on the enthalpy diagram on the saturated vapor curve, and connecting the tie line, then extending it on to intersect with the x_D ordinate 0.995, reading $(M_D)_{min}$ = 596 Btu/lb

2. Operating reflux ratio, L/D

Select $(L/D)_{actual}$ = 10 (L/D)min = 10 (0.012) = 0.12

This is not unusual to select an operating reflux ratio ten, or even fifty times such a low minimum. Selection of a higher reflux ratio will result in a reduction in the number of trays required. As a consequence this becomes a balance of the reduction in trays versus operating and capital expense in handling the increased liquid both external to the column and internally.

3. Operating M_D

M_D = 649.8 Btu/lb

Locate this value on the diagram and connect it to the feed point, x_F. Extend this line to intersect the bottoms condition ordinate (extended), x_B. In this case, it is impossible to represent the value, x = 0.005, accurately, but construct it as close as possible to the required condition. M_B is now located. The accuracy of this point will be improved later in the problem.

Following the procedures shown in Figure 10-82A–CFigure 10-88 and Figure 10-89, the trays are constructed from the top or overhead down toward the bottom. The x values are read to correspond to the y values constructed. This establishes the tie line. When the x value tie line points (representing the trays) cross the feed ordinate, the construction is shifted from using the point M_D to the point M_B. Note that only 1½ theoretical trays are required above the feed, since this is predominantly a stripping operation. The number of theoretical trays or stages which can be easily plotted is six to seven counting down from the top. The sixth tray is too inaccurate to use graphically. Instead of calculating the balance of the trays assuming a straight line equilibrium equation from tray seven to the end, the plot could be enlarged in this area and the trays stepped off. By reference to the x–y diagram it can be seen that the equilibrium line from x = 0.02 to x = 0 is straight.

For the condition of straight operating and equilibrium curves, the number of plates can be calculated including the “reference” plate (number seven in this case) [59].

(10-414)

(10-414)

where

N_B = number of trays from tray m to bottom tray, but not including the still or reboiler

x_m = tray liquid mol fraction for start of calculations (most volatile component)

x_{1B} = mol fraction most volatile component in bottoms

For the lower end of the equilibrium curve,

$y_m = 5.0\ x_m$ (by slope calculation of x–y diagram)

For the stripping section: consider top seven trays, vapor entering tray No.6, $y_7 = 0.02$, m = tray 7, m + 1 = tray 6, reading from diagram,

use $H_m = 1{,}190$

$h_{m+1} = 369$

$M_B = -960$

N_B = 1.71 trays (theoretical) not including reboiler, but including tray number 7, the one used as reference.

Total trays = 7 (from diagram plus (1.71 − 1) = 7.7 theoretical, plus a reboiler or 8.7 including a reboiler.

Tray efficiency is calculated as previously described, but stripping tray efficiencies are normally lower than rectification efficiencies. For ammonia–water stripping, most overall efficiencies are 50–60%.

Note that if the problem of accurate graphical representation occurs in the rectification end of the diagram, the corresponding relationship to use in order to calculate the balance of the trays, (assuming straight line operating and equilibrium lines in the region) is [59]:

Rectifying section:

(10-415)

where

K' = equilibrium constant for the least volatile component, K' = y/x

N_n = number of plates above (but not including) reference plate n

y′, x′ = mol fractions least volatile component

Example 10-39 Ponchon-Savarit Unequal Molal Overflow in heptane-ethyl benzene System

A saturated liquid feed mixture composed of 45 mole % heptane and 55 mole % ethyl benzene is to be fractionated at 760 mm Hg to produce a distillate containing 98 mole % heptane and a residue containing 99 mole % ethyl benzene. Using the enthalpy-concentration diagram based on molal enthalpy and mole fraction as shown in the table below, determine the following:

1. The minimum reflux ratio L_o/D

2. The number of theoretical stages for an operating reflux ratio of 2.5 R_{min}.
3. The condenser duty kcal/kg mole.
4. The reboiler duty, kcal/kg mole.

Solution

Temperature°C	x_H	y_H	H_L, kcal/kg mole	H_V, kcal/kg mole
136.2	0	0	5.92	14.52
129.5	0.08	0.233	5.697	14.203
122.9	0.185	0.428	5.52	13.92
119.7	0.251	0.514	5.44	13.87
116.0	0.335	0.608	5.36	13.61
110.8	0.487	0.729	5.26	13.45
106.2	0.651	0.834	5.17	13.11
103.0	0.788	0.904	5.16	12.94
100.2	0.914	0.963	5.124	12.784
98.5	1.0	1.0	5.1	12.675

Figure 10-90a and Figure 10-90b show the VLE and enthalpy composition diagrams of heptane respectively from the above data.

Figure 10-90a VLE diagram of heptane-ethylbenzene mixture at 760 mm Hg.

Figure 10-90b Enthalpy composition diagram of heptane-ethyl benzene mixture at 760 mm Hg.

On the enthalpy–composition diagram, vertical lines are constructed at $x_B = 0.01$, $x_F = 0.45$ and $x_D = 0.98$. The point F is the intersection of the feed composition line and the liquid enthalpy curve. The point N is located by producing the line FD to the vertical through x_D. The point D on the vapor enthalpy curve corresponds to the value of y in equilibrium with $x_F = 0.45$ and is obtained from the equilibrium curve at y = 0.699.

1. The point N enables the minimum reflux ratio to be determined from:

And the length NB represents the heat removed in the condenser per unit mass of liquid at its boiling point. Points vertically above N correspond to actual reflux ratios (such as P_1), and thus the greater the value of R, the greater is the distance P_1N and fewer number of plates are required for the separation. However, as P_1B represents the condenser duty, this also rises so that an economic selection of the reflux ratio must be made.

= (21.5 − 12.7)/(12.7 − 5.1)
= 1.158

2. The number of theoretical stages for an operating reflux ratio of 2.5 R_{min} is determined as follows: R = 2.5 × 1.158
= 2.895

At the reflux ratio of 2.895, the point P_1 on Figure 10-90b is determined by:

P_1 = 34.70 kcal/kg mole.
Hence P_1 has coordinates (0.98, 34.70).
The point P_2 is located by producing the line P_1F to its intersection with the vertical through x_B. The length P_2C represents the heat input to the reboiler per unit mass of bottom product. The coordinate of P_2 is −19 kcal/kgmole having the coordinate (0.01, −19). Points P_1 and P_2 are known as the difference points on the rectifying and stripping sections of the column. The construction of the tie lines may start from either side of Figure 10-90b, indicating either the conditions at the top or the bottom of the column and proceed as indicated for the rectifying and stripping sections of the column. In either case, when an equilibrium tie line crosses the line connecting the difference points (P_1P_2) through the feed condition, the other difference point is used to complete the construction. The number of tie lines indicates the theoretical number of stages.
From Figure 10-90b, the number of theoretical stages is 11.

3. The heat removed in the condenser per unit mass of product (condenser duty) in kcal/kg mole is determined from Figure 10-90b by length P_1B.

$P_1B = 34.7 - 5.1 = 29.6$ kcal/kg mole.

4. Heat required in the reboiler per unit mass of product is (reboiler duty) in kcal/kg mole is length P_2C in Figure 10-90b.

$P_2C = 5.9 - (-19) = 24.9$ kcal/kg mole.

Example 10-40 Ponchon-Savarit Unequal Molal Overflow in A–B System

A feed stream of A and B consisting of boiling liquid and saturated

vapor whose enthalpy is 1200 kJ/mol contains 40 mol % of A, and is to be separated to produce a distillate of 99% of A and a bottom product containing not more than 2% of A. Calculate:

1. The minimum number of theoretical plates in the column to effect the separation.
2. The minimum reflux ratio, R_{min}.
3. The number of actual plates required if the column efficiency is 60 % and the reflux ratio $1.2R_{min}$.
4. The optimum feed plate for this condition.

Solution

Equilibrium and Enthalpy data

Vapor liquid equilibrium data

x_A	0	0.05	0.10	0.15	0.20	0.30	0.40	0.50	0.60	0.80	1.0
y_A	0	0.30	0.58	0.71	0.79	0.90	0.96	0.98	0.99	0.995	1

Enthalpy data (kJ/kmol)

x_A	0	0.20	0.40	0.60	0.80	0.9	0.99
$H_{L,}$ kJ/kmol	767	418	198	116	163	209	279
$H_{V,}$ kJ/kmol	2765	2580	2370	2160	1905	1765	1490

1. Figure 10-91 shows the enthalpy composition diagram of A, and the minimum theoretical plates occurs at total reflux when the point P_1 on $x = x_D$ is at infinity. The operating lines are therefore vertical. Further, the condition of the feed is located

between the liquid and vapor enthalpy lines for a partly vaporized feed. For a feed at its boiling point, or as a saturated vapor, F lies on the liquid and vapor enthalpy lines respectively. Similarly, for a liquid below its boiling point or as superheated vapor, the point F lies below or above the respective curves. With the composition and percent vapor specified, the point F may be located.

Figure 10-91

Enthalpy vs composition of A–B system.

For the minimum number of theoretical plates, starting at the top plate, the vapor leaving the plate has composition $y_T = 0.99$. The corresponding value of x_T from the equilibrium data is $x_T = 0.6$. From these points, the equilibrium tie line $x_T y_T$ is drawn. At total reflux, the operating line is vertical and $y_{T-1} = 0.6$ and $x_{T-1} = 0.11$. y_{T-2} is located at $y_{T-2} = 0.11$ and x_{T-2} is found to equal $x_{T-2} = 0.02$. Therefore, two theoretical plates are required at total reflux.

2. The minimum reflux ratio is located by producing the tie line which passes through F to the line $x = x_D$ at N.

Then

= (2880 – 1500)/(1500 – 290)

= 1.14

3. $R = 1.2\ R_{min} = 1.368$. Hence the difference points P_1 and P_2 are $P_1 = 3155$ kJ/mol, $P_2 = -80$ kJ/mol. The theoretical stages are stepped-off as shown in Figure 10-91 and it is found that five

ideal plates are required. For the column efficiency = 60%, the number of actual plates is 5/0.60 = 8 plates.

Figure 10-91a VLE data of ammonia–water system at 1 atm.

4. The feed tray is located on the diagram by the tie line which passes through the line P_1FP_2. The optimum feed plate is between the 2nd and 3rd stage.

10.37. Multicomponent Distillation

The basic background and understanding of binary distillation applies to a large measure in multicomponent problems. Reference should be made to Figure 10-1 for the symbols.

Multicomponent distillations are more complicated than binary systems due primarily to the actual or potential interaction of one or more components of the multicomponent system on the others. These interactions may be in the form of vapor-liquid equilibria, such as azeotrope formation, chemical reaction, etc., which may affect the activity relations, and hence cause deviations from ideal behavior. For example, some systems are known to have two azeotrope combinations in the distillation column. Sometimes these can be “broken” by addition of a third chemical or hydrocarbon.

To properly handle the changing composition relationships, it is almost essential to use computational techniques if good accuracy is to be achieved. Even calculations three component systems become tedious using calculators without significant internal memory.

Computers can be for programmed to easily handle the complexities of calculations involving techniques of trial and checking for convergence to a preset acceptable limit.

Digitally programmable algorithms for determining convergence are often the heart of an efficient multicomponent calculation. There are several techniques incorporated into many programs [27, 76, 112, 135, 139, 168]. Several process simulators on the market can be readily employed to size multicomponent distillation/fractionation columns and Table 10-32 shows the most commonly used commercial process simulators. Most of these programs can be licensed by colleges/universities for educational purpose at nominal cost.

Table 10-32 Simulation Packages.

Name	Type	Source	Internet address (http//www.)
HYSY	Steady-state and dynamic	Aspen Technology Inc. Ten Canal Park Cambridge, MA 02141-2201, USA	Aspentech.com
Aspen Plus	Steady-state	Aspen Technology Inc. Ten Canal Park Cambridge, MA 012141-2201, USA	Aspentech.com
CHEMCAD	Steady-state	Chemstations, Inc. 2901 Wilcrest Suite 305 Houston, TX 77042	Chemstation.net
PRO/11 and DYNSIM	Steady-state and dynamic	SimSci-Esscor 5760 Fleet Street Suite 100 Carlsbad, CA	Simsci.com
DESIGN II	Steady-state	WinSim Inc. P.O. Box 1885	Winsim.com
UniSim Design	Steady-state and dynamic	Honeywell 300-250 York Stree London, Ontario N6A 6K2, Canada	Honeywell.com

ProSim	Steady-state and dynamic	ProSim SA Stratege Batiment A BP 27210 F-31672 Labege Cedex, France	prosim.net

Note: Contact the companion website to check the full features of the most recent versions of the programs.

Key Components

The two components in a feed mixture whose separations will be specified.

1. Adjacent keys: key components that are adjacent with respect to their volatilities.
2. Split keys: key components that are separated in volatilities by a non-key component, i.e., the system of components contains one or more whose volatilities fall between the volatilities of the designated keys.
3. Light key: the designation of the key component with the highest volatility of the two key components.
4. Heavy key: the designation of the key component with the lowest volatility of the two key components.
5. Example: component designations.

Component	**Relative Volatility $\alpha_{l/h}$−7°F. and 550 psia**	**Designation**
Hydrogen	11.7	Lighter than Key
Methane	3.7, α_l	*Light Key, 1*
Ethylene	1.0, α_h	*Heavy Key, h*
Ethane	0.72	Heavier than Key
Propylene	0.23	Heavier than Key
Propane	0.19	Heavier than Key

Hengstebeck [137] presents a simplified procedure for reducing a multicomponent system to an equivalent binary using “key” components. From this, the number of stages or theoretical plates and reflux can be determined using conventional binary procedures and the McCabeThiele method.

Liddle [136] presents a shortcut technique for multicomponent calculations based on improving the Fenske and Gilliland correlations.

10.37.1. Minimum Reflux Ratio-Infinite Plates

This is the smallest value of external reflux ratio (L/D) which can be used to obtain a specified separation. This is not an operable condition. Knowledge of this minimum reflux ratio helps to establish an economical and practical operating ratio. Ratios of 1.2 to 2.0 times the minimum are often the economical range for hydrocarbon systems. However, as already discussed, high reflux rates increase column size (but reduce number of trays required), reboiler size, steam rate, condenser size and coolant rate.

For adjacent key systems, all components lighter than the light key appear only in the overhead, and all components heavier than the heavy key appear only in the bottoms, and the keys each appear in the overhead and bottoms at levels in accordance with specifications.

For a split key system in the lights and heavies distribute the same way as adjacent key systems. However, the component(s) between the keys also distribute to overhead and bottoms.

At minimum reflux, the regions in which the number of trays approaches infinity (called the pinch zones and region of constant compositions) are:

1. Binary system: pinch zone adjacent to feed plate.
2. Multicomponent:

a. Three components with no component lighter than light key: pinch zone in stripping section adjacent to feed plate
b. Three components with no component heavier than heavy key: pinch zone in rectifying section adjacent to feed plate
c. Three components mixture: pinch zones may be above and below feed plate
d. Greater than four components: pinch zones appear in rectifying and stripping sections

For systems with one sidestream drawoff, either above or below the feed, Tsuo *et al.* [102] propose a method for recognizing that the minimum reflux ratio is greater for a column with sidestream drawoff. At the sidestream the operating line has an inflection. For multifeed distillation systems, the minimum reflux is determined by factoring together the separate effect of each feed [103].

Lesi [105] proposes a detailed graphical procedure for figuring multicomponent minimum reflux by a graphical extension of a McCabeThiele diagram, assuming infinite plates or equilibrium stages. In this traditional model, the concentration in the distillate of the components heavier than the heavy key are assumed to be zero, and the heavy key component reaches its maximum concentration at the upper pinch point (see Figure 10-48 and Figure 10-50). Therefore, the assumption is that only the heavy and light keys are present at the upper pinch point, similar in concept to the handling of a binary mixture [106]. The method assumes (a) only the key components are distributed, (b) no split key components exist, (c) total molal overflow rates and relative volatilities are constant. This method provides good agreement with the detailed method of Underwood.

Yaws *et al.* [124] provide an estimation technique for recovery of

each component in the distillate and bottoms, from multicomponent distillation, using shortcut equations and involving specification of the recovery of each component in the distillate, recovery of the heavy key component in the bottoms, and the relative volatility of the light key component. The results compare very well with plate-to-plate calculations. Figure 10-92, for a wide range of recoveries of 0.05 to 99.93% in the distillate.

Figure 10-92

Yaws shortcut method compared to plate-to-plate calculations. Used by permission, Yaws, C. L. *et al.* Hydrocarbon Processing, V. 58, No. 2 (1979) p. 99. Gulf Publishing Co., all rights reserved.

Equations Surveyed

The availability of desktop computers and simulation software packages has permitted optimization of the number of equilibrium stages in multipurpose fractionating towers and absorbers. Vigorous iterative plate-to-plate computation is still used to study a wide range of process conditions. However, preliminary design with graphical correlations often helps to arrive at an approximate optimum number of stages before reverting to established design methods. Figure 10-1 shows a schematic of a fractionating column with two or more multicomponent and associated equipment items. Alternatives to the preliminary design approach are the shortcut-methods that estimate a realistic optimum number of theoretical stages. Generally, shortcut methods were developed to establish the minimum number of stages at total reflux, and also the minimum reflux at an infinite number of theoretical stages. Many authors have correlated the minimum reflux

and minimum stages with operating reflux and corresponding theoretical equilibrium stages required. These methods have enabled the reflux ratio and number of plates to be chosen to achieve an optimum relationship based on investment and operating costs.

Mathematical Modeling

The distribution of components between the distillate and bottoms is given by the Hengstebeck-Geddes equation [124, 125, 126]:

(10-416)

where

d_i = moles of component i in the distillate

b_i = moles of component i in the bottoms.

α = relative volatility of component i

A, B = correlation constants.

A material balance for the ith component in the feed is

(10-417)

The quantity of component i in the distillate can be expressed as mole fraction recovered, or . Alternatively, in the bottoms, the mole fraction of component i recovered is .

If Equation 10-416 is expressed with respect to the heavy key component, then

(10-418)

The relative volatility of the heavy key component (i.e. $\alpha_{HK} = 1.0$).

Therefore,

(10-419)

but

(10-420)

and

(10-421)

Therefore

(10-422)

Equation 10-422 can be expressed in terms of mole fraction recovered as:

(10-423)

Substituting Equation 10-419 into Equation 10-416 and expressing in terms of the light key component.

(10-424)

Therefore:

(10-425)

Expressing Equation 10-425 in terms of fractional recoveries:

(10-426)

(10-427)

The recoveries of the ith component in the distillate and bottoms:

Equation 10-416 and Equation 10-417 are:

(10-416)

(10-417)

Expressing Equations 10-416 and 10-417 in terms of the recoveries of

the ith component.

That is:

(10-428)

Hence:

(10-429)

(10-430)

The recovery of the ith component in the distillate is given by:

(10-431)

and the recovery of the ith component in the bottoms is:

(10-432)

(10-433)

(10-434)

(10-434)

The correlation constants required for Equations 10-431 and 10-434 are obtained by specifying a desired recovery of the light key component (LK) in the distillate, and the recovery of the heavy key component (HK) in the bottoms. The constants are then calculated as follows:

(10-435)

or

(10-435a)

(10-436)

10.37.2. The Fenske's Method for Total Reflux [285]

Fenske's [285] equation for determining the minimum equilibrium stages at total reflux was based on an ideal mixture. This suggests that the ratio of vapor pressures or the ratio of equilibrium vaporization of the key components is constant over the range of temperatures (that is, the relative volatilities are constant). Fenske expressed the minimum number of equilibrium stages as:

(10-182)

10.37.3. The Gilliland Method for Number of equilibrium Stages [23]

The number of theoretical equilibrium stages required for a given separation at a given reflux ratio is often determined by empirical correlations [23, 286]. The abscissa X represents a reflux function as:

(10-219)

Correspondingly, the ordinate Y represents a stage function as given by

(10-220)

The ratios used for the axes of abscissa and ordinate were chosen because they provide fixed end points for the curve (X = 0.0, Y = 1.0 and at X = 1.0, Y = 0.0). The two functions were found to give good correlations. These correlations are shown in Figure 10-49a and Figure 10-93. Gilliland's correlation has produced relevant results that offer the advantages that:

- They represent an optimum solution concerning the location of the feed plate.
- The splitting for the two key components is verified.
- The maximum deviation using Gilliland's correlation in terms of tray number is within the 7% range [23].

Figure 10-93

Gilliland correlation of reflux ratio, theoretical stages, minimum reflux ratio and minimum number of stages **a** Excel program snapshot for Example 10-41. **b** Excel program snapshot for Example 10-41.

[source: Gilliland, Ind. Eng. Chem., 32: 1101 (1940)]. $(N + 1)_m$ = The number of theoretical stages at total reflux; R_m = The minimum reflux ratio; R = Reflux ratio; N = Number of theoretical stages.

The Gilliland correlation tends to be conservative for feeds with low values of the thermal condition of the feed (q). It can give inaccurate results when there is a large difference in tray requirements above and below the feed. The correlation is adequate for use in preliminary designs before a detailed analysis is carried out, but great caution should be exercised if it is used for final designs. Hines and Maddox [266] have described an improved correlation for determining the number of theoretical stages.

10.37.4. Underwood's Method [72, 73]

If an infinite or nearly infinite number of equilibrium stages is involved, a zone of constant composition must exist in the fractionating column. In this instance, there is no measurable change in the composition of liquid or vapor from stage or stage. Under these conditions, the reflux ratio can be defined as the minimum reflux ratio, R_{min}, with respect to a given separation of two key components (that is, the light key and heavy key) [72].

Therefore, for component i in the distillate:

(10-437)

where θ is Underwood's constant (or root of the equation), and must

lie between the relative volatilities of the light and heavy keys (α_{LK} and α_{HK}). The number of components is n.

Correspondingly, for the feed:

(10-438)

Equation 10-438 shows the relationship for the feed, where q is the fraction of feed that is liquid at the feed tray temperature and pressure. For a bubble point feed, q = 1, and for a dew point feed, q = 0. The minimum reflux ratio is determined from Equation 10-438 by substituting into Equation 10-437. This author [287] has developed a numerical method for computing θ and R_{min} respectively. However, other methods should be tried, if R_{min} gives a negative value. Also, it may be that the separation between the feed and the overhead can be accomplished in less than one equilibrium stage.

10.37.5. Equations for Describing Gilliland's Graph

Many equations have been proposed to describe Gilliland's curve for multicomponent distillation. However, the difficulty with some of these equations has been in meeting the end conditions of X = 0, Y = 1 and X = 1, Y = 0. A review of the many equations proposed by these authors is as follows.

Proposed Equations

1. Hengstebeck [106] (1961).

(10-439)

where

A = −1.3640187
B = −3.0920489
C = −3.407344729
D = −1.74673876
E = −0.33268897

2. Liddle [136] (1968)

(10-440)

(10-441)

(10-442)

3. Van-Winkle and Todd [139] (1971)

(10-443)

(10-444)

4. Molokanov [288] (1972)

(10-445)

5. Hohman and Lockhart [289] (1972)

(10-446)

6. Eduljee [84] (1975)

(10-447)

7. Huan-Yang Chang [290] (1981)

(10-448)

8. Harg [291] (1985)

(10-449)

9. McCormick [97] (1988)

where

(10-450)

10. Rusche [292] (1999)

(10-451)

where

(10-452)

from which the number of theoretical trays is:

(10-453)

The Gilliland correlation is conservative for feeds with low values of q (thermal condition of the feed), and can result in erroneous results when there is large difference in tray requirements above and below the feed. The various correlations illustrated are for preliminary exploration of design variables, which should be further refined using simulation packages (Aspen, Chemcad, UniSim® etc.). These correlations should be employed with caution, if applied to final design.

From the equations listed, those of McCormick give good agreement in the normal operating range of real towers. The reflux ratio, R is calculated as a multiple of the minimum reflux ratio, R_{min}.

That is:

(10-454)

10.37.6. Operating reflux ratio, R

As the reflux ratio increases above the minimum, the number of trays required for a given separation decreases, hence the capital cost of the column decreases. However, increasing the reflux ratio will increase the vapor rate in the tower, and this will impact the cost of the reboilers and condensers, as well as increasing cooling water and steam costs. Hence there is an optimum reflux ratio for any specified separation as illustrated later in this volume. Experience has shown that the value of this optimum normally falls in the range 1.03 <

$R/R_{min} < 1.3$. For the first estimate, the multiplier "FACTOR" generally varies from 1.2 to about 1.5 for conventional columns, but because of economics, the range is now between 1.05 and 1.20.

Rule of thumb:

The operating reflux ratio is chosen so that $R/R_{min} = 1.2$

10.37.7. Feed Tray Location

The approximate location can be determined by the ratio of the total number of theoretical stages above and below the feed plate from the Fenske total reflux relation:

(10-455)

The relation is solved for S_r/S_s. The results are not exact, because the feed tray composition is very seldom the same as the feed; which is the assumption in this relation. Actually, the feed point or correct location for the feed may be wrong by two or three theoretical trays. This will vary with the system. It does mean, however, that when this approach is used for feed plate location, alternative feed nozzles should be installed on the column to allow for location of the best feed point to be determined by experiment. These extra nozzles are usually placed on alternate trays (or more) both above and below the calculated location. A minimum of three alternative nozzles should be available.

When the feed point is located by tray-by-tray calculation, the correct point can be established with greater confidence, but still alternative nozzles are suggested, since even these detailed calculations can be in accurate.

The actual number of trays in the rectifying section $(N_{act})_r$ can be

determined by:

(10-456)

(10-457)

Solve for S_s, because S_M and S_r/S_s are known.

Obtain S_r by difference.

$(N_{act})_r = S_r/E_o$ (for total condenser; if partial condenser use

(10-458)

(10-459)

For systems with wide variation in relative volatility, the suggestion of Cicalese *et al.* [9] is often used to evaluate the theoretical total equilibrium stages in the rectifying and stripping sections:

(10-460)

(10-461)

where

S_r = number theoretical trays/plates in rectifying section

S_s = number theoretical trays/plates in stripping section

Maas [108] presents a useful analysis for selecting the feed tray in a multicomponent column. For accuracy it involves the use of a tray-by-tray computer calculation.

10.37.8. Kirkbride's Feed Plate Location [174]

After the minimum number of stages and the minimum reflux ratio have been determined, the number of theoretical stages is then calculated. The ratio of the number of plates above the feed stage (including the partial condenser) to the number below the feed stage (including the reboiler) can be obtained using Kirkbride's empirical equation [174]. The equation was developed on the basis that the ratio of rectifying trays to stripping trays depends on:

- The fraction of the heavy key component (in the feed) removed in the overhead.
- The fraction of the light key component removed in the bottoms.
- The concentration of the heavy key component in the overhead.
- The concentration of the light key component in the bottoms.

Kirkbride's feed plate equation is expressed as:

(10-462)

or

(10-462a)

Akashah *et al.* [311] presented a modified version of Equation 10-462 or 10-462a, which gives:

(10-462b)

(10-462c)

where

B = molar flow of bottoms product
D = molar flow of top product
N = total number of theoretical stages
m = number of theoretical stages above the feed plate, including any partial condenser
p = number of theoretical stages below the feed plate, including the reboiler
$(x_{LK})_F$ = concentration of the light key in the feed.
$(x_{HK})_F$ = concentration of the heavy key in the feed.
$(x_{LK})_B$ = concentration of the light key in the bottoms product.
$(x_{HK})_D$ = concentration of the heavy key in the distillate product.

Example 10-41: Multicomponent Distillation by Yaw's Method [124] (used by permission)

Assume a multicomponent distillation operation has a feed whose component concentration and component relative volatilities (at the average column conditions) are as shown in Table 10-33. The desired recovery of the light key component O in the distillate is 94.84%. The recovery of the heavy key component P in the bottoms is 95.39%.

Table 10-33 Yaw's Method for Selected Distillation Recovery from a Specific Feed for Example 10-41.

Component	f_i	α_i
M	0.10	2.30
N	0.13	1.75
O (LK)	0.25	1.45
P (HK)	0.23	1.00
Q	0.15	0.90
R	0.08	0.83
S	0.06	0.65
Used by permission, *Hydrocarbon Processing*, Yaw's, C. L., *et al.* V. 58 No. 2 (1979), p. 99, Gulf Pub. Co., all rights reserved.		

The recoveries of the non-key components are estimated by first calculating the correlation constants:

$b_{HK}/f_{HK} = 0.9539$, given

$d_{LK}/f_{LK} = 0.9484$, given

$\alpha_{LK} = 1.45$, from Table 10-33.

The recovery of component M in the distillate is then $d_M/f_M = (10^{-1.3158}2.30^{15.988})/(1 + 10^{-1.3158}\ 2.30^{15.988}) = 0.99997$, from Equation

10-431.

The recovery of component M in the bottoms is $b_M/f_M = 1/(1 + 10^{-1.3158}2.30^{15.988}) = 0.00003$, from Equation 10-434.

Repeating Equations 10-431 and 10-434 for each of the other non-key components in the feed mixture gives the results shown in Table 10-34. Good agreement was demonstrated.

Table 10-34 Results for Example 10-41 for Multicomponent Distillation.

Component	Percent recovery	
	In dist. (100 di/fi)	**In btms. (100 di/fi)**
M	99.997	0.003
N	97.731	2.269
O (LK)	94.840	5.160
P (HK)	4.610	95.390
Q	0.889	99.111
R	0.245	99.755
S	0.005	99.995
Used by permission, *Hydrocarbon Processing*, Yaw's, C. L., et al V. 58 No. 2 (1979), p. 99, Gulf Pub. Co., all rights reserved.		

Computer Program

The author [287] has developed a program for shortcut multicomponent distillation calculations based on a series of equations developed by Hengstebeck-Geddes [124, 125, 126, 137], while the minimum tray calculations (at total reflux) use the equations developed by Fenske [285]. The number of theoretical equilibrium stages is determined using Gilliland's [23] empirical correlations. The program has an option of entering either the equilibrium constant or the relative volatility with the feed-stream key components.

It also arranges the relative volatilities in order of decreasing magnitude before proceeding to compute the correlation constants. The following assumptions are used in the program.

- There is only one feed stream with 12 or fewer key components.
- There is only one heavy key component.
- Feed components may be arranged in order of decreasing relative volatility, but the light key (LK) and heavy key (HK) components must be adjacent to each other.
- The feed enters the column at the optimum stage.
- The column produces two products (distillate and bottoms) with overhead condenser and bottom reboiler.

The program uses the bisection method to determine Underwood's constant θ, or the root of Underwood's equation, and then proceeds to evaluate the minimum reflux ratio. From a given multiplier (FACTOR), the actual reflux ratio is calculated.

From the various equations that have been proposed to describe Gilliland's curve for multicomponent distillation, this author uses that of McCormick (Equation 10-450) because it gives a good agreement in the normal operating range of real towers. The program then proceeds to determine the actual reflux ratio from the minimum reflux ratio using Equation 10-454. Using these equations with Equation 10-462, the program calculates the number of theoretical plates and the optimum location of the feed plate in a distillation column.

NB: This program is best used in preliminary design before a detailed design by a process simulator package is explored.

Table 10-35 shows the input data and results of Example 10-41 using

PROG104 is the Absoft Microsoft Runtime Windows Environment (MRWE) Fortran Compiler. An Excel spreadsheet (Example 10-41.xls) has been developed to determine θ, the root of Underwood's equation and the minimum reflux ratio, R_{min} for multicomponent distillation. The program uses Microsoft Excel optimization technique such as Go-Seek or Solver to determine the Underwood constant θ for a given feed condition (q), and then proceeds to calculate the minimum reflux ratio for various values of the distillate composition. The actual reflux ratio is determined by multiplying the minimum reflux ratio by a factor (FACTOR) of 1.15 to 1.3. The program uses the Fenske equation to determine the minimum number of plates at total reflux, using the developed correlation of Erbar-Maddox (Figure 10-94 and Figure 10-94a). The program further calculates the number of theoretical plates and compares the value with the Rusche correlation (Equation 10-451). Figures 10-93a and 10-93b show snapshots of the calculations. The results in Table 10-34 by Yaws et al [124] and in Table 10-35 from the developed computer program PROG104 are in good agreement.

Table 10-35 Input Data and Computer Results for Multicomponent Fractionation

Data Name: Data104.Dat	
Rel	
7	
0.1	2.3
0.13	1.75
0.25	1.45
0.23	1.0
0.15	0.90
0.08	0.83

0.06	0.65
0.9484	0.9539
41	1.20

Multicomponent System Fractionation

Component Number	Feed Moles	Rel Volatility Alpha	Distilattle %	Moles	Bottoms %	Moles
1	0.1000	2.3000	99.9966	0.1000	0.0034	0.0000
2	0.1300	1.7500	99.7317	0.1297	0.2683	0.0003
3	0.2500	1.4500	94.8400	0.2371	5.1600	0.0129
4	0.2300	1.0000	4.6100	0.0106	95.3900	0.2194
5	0.1500	0.9000	0.8886	0.0013	99.1114	0.1487
6	0.0800	0.8300	0.2451	0.0002	99.7549	0.0798
7	0.0600	0.6500	0.0049	0.0000	99.9951	0.0600

Component Number	Feed Moles	Mol Fract. XF	Distilattle Moles	Mol Fract. XD	Bottoms Moles	Mol. XB
1	0.1000	0.1000	0.1000	0.2088	0.0000	0.0000
2	0.1300	0.1300	0.1297	0.2707	0.0003	0.0007
3	0.2500	0.2500	0.2371	0.4951	0.0129	0.0248
4	0.2300	0.2300	0.0106	0.0221	0.2194	0.4210
5	0.1500	0.1500	0.0013	0.0028	0.1487	0.2853
6	0.0800	0.0800	0.0002	0.0004	0.0798	0.1531
7	0.0600	0.0600	0.0000	0.0000	0.0600	0.1151

The EED condition, Q:	1.0
The heavy key component number is:	4
Percentage recovey of the light key component	94.84
In the distillate is (%):	95.39
Percentage recovery of	

Percentage recovery of the heavy key compnent	
In the bottoms is (%) :	0.4789
Total moles in the distillate:	0.5211
Total moles in the bottoms:	1.0000
Total moles in the feed:	1.1850
Underwood constant:	
Factor for the reflux ratio:	1.2000
Minimum reflux ratio:	2.8490
Actual reflux ratio:	3.4188
Minimum number of stages is:	16.0
Number of theoretical plates in the column:	33.6
The position of the feed plate is:	17.2

Figure 10-94

Plate-reflux correlation of Erbar and Maddox.

Figure 10-94a Erbar-Gray–Maddox correlation. Dashed Lines: extrapolated

(source: Erbar and Maddox, Hydrocarbon Process. Petrol Refiner, 4, No. 5, 183, 1961).

10.37.9. Algebraic Plate-to-Plate Method

Like any plate-to-plate calculation, this is tedious and in most instances does not justify the time spent, because shorter methods give reasonably acceptable results. Van Winkle [75] outlines the steps necessary for such calculations.

With current computer technology there are several commercial programs available (as well as personal and private ones) that perform tray-to-tray stepwise calculations up or down a column, using the latest vapor pressure, K-values, and heat data for the components. This then provides an accurate analysis at each tray (liquid and vapor analysis) and also the heat duty of the bottoms reboiler and overhead total or partial condenser.

Torres-Marchal ([110] and [111]) present a detailed graphical solution for multicomponent ternary systems that can be useful to establish the important parameters prior to undertaking a more rigorous solution with a computer program. This technique can be used for azeotropic mixtures, close-boiling mixtures and similar situations.

An alternative improved solution for Underwood's method is given by Erbar, Joyner, and Maddox [113] with an example.

10.37.10. Erbar-Maddox Method [115]

This method is used to determine the actual number of theoretical stages at a practical reflux ratio. It uses a plot of $R/(R + 1)$ against N_{min}/N with $R_{min}/(R_{min} + 1)$ as the parameter. When $R = R_{min}$, the axis becomes zero. Therefore, the y-axis of the diagram represents minimum reflux conditions. When $N = N_{min}$, both x and y coordinates

become unity. Branan [284] has developed a polynomial expression that represents the Erbar-Maddox curves. He represents the equation by the following:

(10-463)

where

A, B, C, D, E, F = constants for varying values of z.

(10-464)

(10-465)

(10-466)

Table 10-36 shows the constant values at varying z. The author has used the values in Table 10-36 and developed an Excel spreadsheet to create the Erbar-Maddox correlation. Figure 10-94 shows the plates-reflux correlation of Erbar-Maddox and is based on a bubble-point feed. The correlations from Figure 10-94 are in good agreement with the Erbar-Maddox correlation of Figure 10-94a. For other types of feed, the following correlation is used [112]:

(10-467)

where

D = Distillate flow rate, lb-mole/h.

F = Feed flow rate, lb-mole/h

H = Enthalpy, Btu/h

L_o = Liquid condensed at the condenser, lb-mole/h.

Q_c = Condenser duty, Btu/h

V = Vapor flow rate in the rectifying section, lb-mole/h.

Table 10-36 Constants for the Erbar-Maddox correlation.

z	A	B	C	D	E	F
0	0.00035	0.16287	−0.23193	5.09032	−8.50815	4.48718
0.1	0.09881	0.32725	−2.57575	10.20104	−12.8205	5.76923
0.2	0.19970	0.14236	−0.58646	2.60561	−3.12499	1.76282
0.3	0.29984	0.09393	−0.23913	1.49008	−2.4388	1.79486
0.4	0.40026	0.12494	−0.49585	2.15836	−3.27068	2.08333
0.5	0.50049	−0.03058	0.81585	−2.61655	3.61305	−1.28205
0.6	0.60063	−0.00792	0.60063	−2.06912	3.39816	−1.52243
0.7	0.70023	−0.01109	0.45388	−1.25263	1.94348	−0.83334
0.8	0.80013	−0.01248	0.76154	−2.72399	3.85707	−1.68269
0.9	0.89947	0.00420	0.38713	−1.14962	1.40297	−0.54487
1.0	1.0	0	0	0	0	0

Source: Branan, C.R., Rules of Thumb for Chemical Engineers: A manual of quick, accurate solutions to everyday engineering problems, 4th Ed., Elsevier, 2005.

Subscripts

F = feed.

k = at the bubble point.

u = at the conditions differing from bubble point.

In the Erbar-Maddox correlation, minimum stages are calculated by the Winn method, and minimum reflux by the Underwood method-but the Fenske minimum stages method can also be used.

10.37.11. Underwood Algebraic Method: Adjacent Key Systems [72]

This system for evaluating multicomponent adjacent key systems, assuming constant relative volatility and constant molal overflow, has proven generally satisfactory for many chemical and hydrocarbon applications. It gives a rigorous solution for constant molal overflow and volatility, and acceptable results for most cases which deviate from these limitations.

Overall Column–Constant α

(10-468)

In arriving at $(L/D)_{min}$ the correct value of θ is obtained from:

(10-469)

The “q” value that is previously described for the thermal condition of the feed.

Rectifying section only:

(10-470)

Stripping section only:

(10-471)

At the minimum reflux condition all the θ values are equal, and generally related:

$\alpha_h < \theta < \alpha_l$

Suggested Procedure

1. From Equation 10-468 expressing θ and q evaluate θ by trial and error, noting that θ will have a value between the α of the heavy key and the α of the light key evaluated at or near pinch temperatures, or at α_{avg}. Suggested tabulation, starting with an assumed θ value, θ_a:

Component	x_{Fi}	$\alpha_i x_{Fi}$	$\alpha_i - \theta$	$\alpha_i x_{Fi}/(\alpha_i - \theta)$	α $x_{Fi}/$ $- \theta$
a	x_{Fa}	$\alpha_a x_{Fa}$	$\alpha_a - \theta_a$	$\alpha_a x_{Fa}/(\alpha_a - \theta_a)$	α_a $x_{Fa}/$ $- \theta$
b	x_{Fb}	$\alpha_b x_{Fb}$	$\alpha_b - \theta_a$	$\alpha_b x_{Fb}/(\alpha_b - \theta_a)$	α_b $x_{Fb}/$ $- \theta$
•	•	•	•	•	•
•	•	•	•	•	•
Ψ, Ψ′, represents function.					

Corrected θ by Newton's approximation method:

(10-472)

Repeat the same type of tabular computation, substituting the corrected θ_c for the θ_a. If the second, corrected θ'_c checks closely with the θ_c, the right value of θ has been obtained, if not, a third recalculation should be made using the θ'_c value as the new assumed value.

Note that average α values should be used (constant) for each component, unless the values vary considerably through the column. In this latter case follow the discussion given elsewhere in this section.

2. Calculate $(L/D)_{min}$ by substituting the final θ value in Equation 10-468. Note that this requires evaluating the functions associated with θ at the composition of the distillate product. The α values are the constant values previously used above.

10.37.12. Underwood Algebraic Method: Adjacent Key Systems; Variable α

For varying α systems, the following procedure is suggested:

1. Assume $(L/D)_{min}$ and determine the pinch temperature by Colburn's method.
2. At this temperature, evaluate α at pinch and α at overhead temperature, obtaining a geometric average α. As an alternative, Shiras *et al.* [63] indicate a t_{avg} value which gives acceptable results when compared to pinch and stepwise calculations. This suggestion calculates:

3. Determine Underwood's θ value as previously described, using the average α value.
4. Calculate $(L/D)_{min}$ and compare with the assumed value of (1) above. If check is satisfactory, $(L/D)_{min}$ is complete; if not,

reassume new $(L/D)_{min}$ using calculated value as basis, and repeat (1) through (4) until satisfactory check is obtained.

where

t_o = overhead temp °F

t_B = bottoms temp °F

t_{avg} = avg temp, °F

To aid in solving the tedious Underwood equation to ultimately arrive at $(L/D)_{min}$, Frank [100] has developed Figure 10-95, which applies for liquid feed at its bubble point and to binary or multicomponent systems, but does require that the key components are adjacent. Otherwise, the system must be solved for two values of θ [74]. To obtain the necessary parameters for Figure 10-95, Frank recommends using the same overhead concentrations that were used in, or calculated by, the Fenske equation for the Underwood solution. (θ = Underwood constant.)

Figure 10-95

Shortcut solution of Fenske-Underwood-Gilliland theoretical trays for multicomponent distillation.

Used by permission, Frank, O., Chem. Eng., Mar. 14, p. 109 (1977).

Example 10-42

An equimolar binary mixture with a relative volatility of 2 is fed to a distillation column which produces a distillate containing 97.5% of the more volatile component and a residue containing 10% of the

more volatile component. The feed is liquid at its boiling point and 80% of the liquid entering the reboiler is vaporized and returned to the column as boil-up.

(i) Estimate the number of theoretical plates provided by this column.

(ii) Check your answer in part (i) by reference to the graph of the Gilliland correlation and Erbar and Maddox correlation.

Solution

The overall mass balance on the column.

(1)

The component balance on the more volatile component (MVC) is:

(2)

Assuming that F = 100 mols/h.

$x_F = 0.5$, $x_D = 0.975$ and $x_B = 0.1$

Substituting the values of the components in Eq. 2 gives:

(3)

Rearranging Eq. 1 and substituting into Eq. 3 gives:

(4)

(5)

D = 45.71 mols/h.

B = 54.29 mols/h.

The upper operating line (UOL) at the top of the column is:

The overall mass balance on UOL is:

(6)

Component balance is:

(7)

or

(8)

(9)

Substituting Eq. 9 into Eq. 6 gives:

(10)

And Equation 8 (UOL), assuming constant molar overflow, becomes:

(11)

The intercept on the UOL, is:

(12)

Since the feed is liquid at its boiling point (i.e. q = 1) and 80% of the liquid entering the reboiler (L′) is vaporized to give the vapor (V′), which leaves the reboiler and returns to the column.

Therefore at the bottom of the column, if L′ = 1.0 and V′ = 0.8.

The mass balance at the bottom of the column with the reboiler is:

(6)

or

(7)

The component balance at the bottom is:

or

where = gradient of lower operating line (LOL). If α = 2, using the Excel spreadsheet (Example 10-42.xls) to plot the VLE diagram, The McCabeThiele diagram of Figure 10-96 is drawn as follows:

Figure 10-96

The McCabeThiele diagram for Example 10-42.

At the coordinates (x_w, x_w) = 0.1 on the 45° line, draw the Lower operating line (LOL) with the slope of 1.25, to intercept the "q" (i.e. q = 1 @ boiling point) line at x_F = 0.5.

At coordinates (x_D , x_D) = 0.975 on the 45° line, draw the upper operating line to intersect the lower operating line (LOL) at q = 1 and extend to the intercept I on the y-ordinate of the VLE diagram.

Step-off the line between the UOL, LOL and the 45° line from x_D = 0.975 to x_W = 0.1 to determine the number of equilibrium stages.

R = 3.43

From McCabeThiele diagram (Figure 10-96):

N +1 = 14 stages

N = 13 stages

(b) R_{min}: from the points (x_D, x_D), a straight line is drawn that touches the VLE line with the "q" line and extended to the y-axis at a point .

Using the Underwood Equation for the minimum reflux ratio:

(10-210)

The Fenske method for the minimum number of theoretical stages at total reflux is:

(10-182)

From the Gilliland correlation (Figure 10-93), the abscissa axis is:

(10-219)

The corresponding point on the ordinate axis of the Gilliland correlation is 0.31

Therefore the number of theoretical stages can be determined from the ordinate by:

(10-220)

N = 12.69 stages (13 stages)

Using the Erbar and Maddox correlation (Figure 10-94 and Figure 10-94a).

The lines of constant is

1.85/2.85 = 0.649 (0.65)

The ordinate axis of Erbar and Maddox correlation is:

A line is drawn at point 0.79 on the ordinate of the correlation and intercepts the lines of constant at 0.65. The corresponding point on the abscissa is 0.74

The number of theoretical stages is:

N = 8.45/0.74 = 11.4 stages.

Example 10-43. Using Data from Example 10-20 (Erbar-Maddox correlation Figure 10-94a)

Using the Erbar and Maddox correlation.

The lines of constant is

1.85/2.85 = 0.649 (0.65)

The ordinate axis of Erbar and Maddox correlation is:

A line is drawn at point 0.79 on the ordinate of the correlation and intercepts the lines of constant at 0.65. The corresponding point on the abscissa is 0.74.

The number of theoretical stages is:

N = 8.45/0.74 = 11.4 stages.

10.37.13. Underwood Algebraic Method: Split Key Systems: Constant Volatility [72]

This method appears tedious, but it is not so unwieldy as to be impractical, but it does require close attention to detail. However, a value of $(L/D)_{min}$ can be obtained with one trial that may be satisfactory for "order of magnitude" use, which is often all that is needed before proceeding with detailed column design and establishment of an operational L/D ratio.

1. Assume θ_f values and check by:

(10-473)

There are total solutions of θ_{fi} equal to one more than the number of split components between the keys. The θ_f values will be spaced:

where α_l is the light key and component number 3, and

correspondingly for the heavy key, component number 6. Determine θ values as for constant volatility case of adjacent keys.

For some systems, the θ values can be assumed without further solution of the above relation, but using these assumed values as below.

2. Calculate:

(10-474)

which represents (for the hypothetical system set up in (1)) the product (θ_{f5}) (θ_{f4}) (θ_{f3}) divided by the product of (α_5) (α_4), based upon the lightest component being numbered one, the next two, etc., the heaviest components having the higher numbered subscripts. P means product, and 1, i = h − 1, i = 1 + 1 are limits for evaluation referring to components between the keys, and the light and heavy keys.

3. Calculate:

(10-475)

For the θ example shown in (1) above:

Also calculate ω for all components lighter than light key.

Component,	ω_j	α_j		

J				
L (light key)	•	•	•	•
H (heavy key)	•	•	•	•
L_{l+1} light				
L_{l+2} than light	•	•	•	•
L_{l+3} key, *etc.*				

4. Calculate $(L/V)_{min}$: (internal)

(10-476)

5. Calculate External $(L/D)_{min}$:

(10-477)

For variable α conditions, the pinch temperature can be used for α determinations as previously described.

Example 10-44: Minimum Reflux Ratio Using Underwood Equation: Calculate the Minimum Reflux Ratio

Use ϕ_{fl} = 0.584 to begin, (assumed).

Expanding to determine more exact value of ϕ_{fl}:

Component	x_{fi}	$\alpha_i x_{fi}$	$\alpha_i - \phi_a$	
1	0.10	0.025	−0.334	−0.0749
2	0.225	0.1125	−0.084	−1.34
3	0.450	0.450	+0.416	+1.08
4	0.225	0.450	+1.416	+0.318
				$\sum = -0.016$
			$\Omega\phi_a = -0.016 - (1 - q)$	
			$\phi_a = -0.016 - (1 - 1)$	
			$\phi_a = -0.016$	

$\Phi_{fl\text{-corrected}} = 0.584 - (-0.016)/(18.948) = 0.584 + 0.00084 = 0.58484$

Now use this new value of Φ_{fl} in Underwood's equations;

For minimum reflux: $\Phi_{fl} = \Phi_r = \Phi_s$

From calculations of related problem* the value of Dx_{Di} has been calculated:

$Dx_{Di} = 0.01072$ for heavy key

Dx_{Di} = 0.428 for light key
Dx_{Di} = 0.225 for lighter than light key

$$[-0.638 + 1.03 + 0.318] - V_r = 0$$

From related problem, D has been determined to be = 0.67 mols/mol feed.

$$-V_r = -1.285$$

$$V_r = 1.285$$

$$(V_r)_{min} = (L_r)_{min} + D$$

$$(L_r)_{min} = 1.285 - 0.6637 = 0.622$$

, Minimum Reflux Ratio

10.37.14. Minimum Reflux Colburn Method: Pinch Temperatures [12]

This method has also found wide usage and might be considered less tedious by some designers. It also yields an approximation of the rectifying and stripping section pinch temperatures. For adjacent keys,

Rectifying:

(10-478)

where

α = relative volatility of any component referenced to the heavy

key component

x_{hD} = overhead composition of heavy key component, mol frac

x_{hn} = pinch composition of heavy key component, mol frac

x_D = overhead composition of any light component, mol frac

x_n = pinch composition of any light component, mol frac

1. Calculate D, B, Dx_{Di} and Bx_{Bi} from problem specification.

2. Assume or set the operating pressure and overhead temperature (may be calculated).

3. Calculate the liquid and vapor quantities and their respective compositions in the feed to the column.

4. Calculate estimated ratio of key components on feed plate based on the liquid portion of the feed.

a. For all liquid feed at feed tray temperature (boiling point) r_f = mol fraction ratios in feed.

b. For a part or all vapor feed just at its dew point, r_f = ratio of key components in the equilibrium liquid phase of feed.

c. For all liquid feed below feed plate temperature r_f = ratio of key components at intersection point of operating line (from a McCabeThiele diagram).

5. Determine approximate pinch zone liquid composition for light key component

(10-479)

all components in liquid portion of feed heavier than heavy key. Note that x_{Fi} values are the mol fractions of the component in the liquid portion of feed only and the $\sum x_{Fi}$ equal to 1.0.

6. Calculate approximate value for $(L/D)_{min}$.

The second term in the right hand parentheses can be omitted unless the mol fraction of the heavy key in the distillate, x_{hD} is 0.1 or greater. Use $x_{hn} = x_{ln}/r_f$.

7. Estimate stripping and rectifying pinch temperatures at values one-third and two-thirds of the interval between the column bottoms and overhead, respectively.

8. Calculate internal vapor and liquid flows.

$(L_r/D)_{min}$ = assumed

Solve for $(L_r/V_r)_{min}$

(10-480)

L_r = (number) (V_r) = $(L_r/D)_{min}(D)$ D is known

Calculate V_r and L_r from above.

In stripping section:

Solve directly for L_s

$L_s = L_r + qF$

Solve for V_s:

(10-481)

Calculate L_s/B

9. Evaluate pinch compositions at the assumed temperatures of

Step 7. If this temperature does not give a balance, other temperatures should be assumed and a balance sought as indicated below. Either of the following balances can be used, for convenience: Rectifying:

(10-482)

When the heavy key in the overhead is very small, less than 0.1 mol fraction, the last term of the denominator can be omitted.

Note that the calculations are only made for the heavy key, h; light key, 1; and all components lighter than it, L′. If there are split keys, the calculation is to include all components lighter than the heavy key.

Stripping pinch compositions:

(10-483)

Because the second term of the denominator is usually negligible when the light key in the bottoms is very small; (less than 0.1 mol fraction), this term is often omitted.

Note that these calculations are made for the light key, 1; heavy key, h; and all components heavier than the heavy key, H. For split key systems, the calculations are made for all components heavier than the light key.

10. Calculate mol fraction ratio:

a. Stripping pinch.

b. Rectifying pinch

c. $p = r_{ps}/r_{pr}$

11. Calculate for each component in pinch.

Rectifying: apply only to components lighter than light key, i = L'

Read from Figure 10-97 value of C_{ni} for each component.

Figure 10-97

Colburn minimum reflux factors, above (Cn) and (Cm) feed point.

Used by permission, Colburn, A. P., The American Institute of Chemical Engineers, Trans, Amer. Inst. Chem. Engr. Vol. 37, p. 805, (1941). All rights reserved.

Calculate for each component:

Sum these values:

Stripping: apply only to components heavier than heavy keys, i = H.

$(\alpha_1 - 1)\ (\alpha_1)$

Read from Figure 10-97 value of C_{mi} for each component.

Calculate for each component:

Sum these values:

12. Calculate:

(10-481)

If the two values of p are not very nearly equal, this requires a retrial with a new $(L/D)_{min}$, and a follow through of the steps above.

When $r_{ps}/r_{pr} > p'$, the assumed $(L/D)_{min}$ is too high. Note that r_{ps}/r_{pr} changes rapidly with small changes in $(L/D)_{min}$, p' changes slightly. When $p = p'$, the proper $(L/D)_{min}$ has been found. Colburn reports the method is accurate to 1%. It is convenient to graph the assumed $(L/D)_{min}$ versus p and p' in order to facilitate the selection of the correct $(L/D)_{min}$.

Example 10-45: Application of the Colburn Equation to Calculate the Minimum Reflux Ratio for a Multicomponent Mixture

A mixture of four components is as listed below, using n-butane as the base component.

Component	**Relative Vol.**	x_f	$(S_r)_i = Dx_D/Bx_B$
1	0.25	0.10	–
Heavy Key	0.50	0.225	0.05
n-butane	1.0	0.450	20.00
4	2.0	0.225	–

S_r = separation ratio =

If all at top, $S_r = 1$

If all at bottom, $S_r = 0$

For component No.3; Basis; 1 mol feed,

For component No. 2,

Because this is at minimum reflux, and adjacent keys system,

$Fx_{FL} = Dx_{DL}$

$Fx_{FH} = Bx_{BH}$

Therefore, for component No.4, lighter than light key,

$Fx_{F4} = (1)\ (0.225) = 0.225$

Then, $Dx_{D4} = 0.225$

For component No.1, heavier than heavy key, this component will not appear in the overhead.

Bottoms:

$Fx_{Fi} = Dx_{Di} + Bx_{Bi}$

, by definition

Then,

Component No. 1:

$Fx_{Fi} = Bx_{Bi} = (1)\ (0.1) = 0.10$

Component No. 2:

Substituting in equation previously established,

or, because Dx_{D2} has been calculated,

Component No. 3:

Component No. 4: This component will not be in the bottoms because it is lighter than the light key:

Overhead:

$Dx_{D1} + Dx_{D2} + Dx_{D3} + Dx_{D4} = D$

$0 + 0.01072 + 0.428 + 0.225 = D$

D = 0.66372 mols overhead product/mol feed

Composition of Overhead:

Component	Dx_{Di}	Mols%
1	0.0	0
2	0.01072	1.6
3	0.428	64.6
4	0.225	33.9
Total	0.66372	100.1

Composition of Bottoms:

Component	Bx_{Bi}	Mols%
1	0.10	29.9
2	0.214	63.8
3	0.0214	6.3
4	0.0	0
Total	0.3354	100.0

To have some idea of what value to use in Colburn's “exact” method for minimum reflux, use Colburn's “approximate” method to establish the order-of-magnitude of the minimum reflux:

where

x_D and x_n = top and pinch compositions of a given light component

x_{hD} and x_{hn} = top and pinch compositions of the heavy key component

α = relative volatility of the given component with reference to the heavy key

Estimating Pinch Composition:

where

r_f = ratio of liquid composition of light to heavy key component on feed plate

x_n = mol fraction of a component in the liquid in the rectifying column pinch

x_z = mol fraction of a component in the liquid part of the feed where the feed is part vapor

$r_f = 0.450/0.225 = 2.0$

$\sum \alpha x_z = (0.5)\ (0.10) = 0.05$

In terms of heavy key:

$\alpha_{l/h} = 2.0$

$\alpha_{l/l} = 0.25$

$\alpha_{l/h} = (\alpha_{l/h})\ (\alpha_{l/l}) = (2.0)\ (0.25) = 0.5$

$x_{D3} = 0.646$

approx.

Now: Use Colburn's more detailed method:

V = L + D

$L_r = (0.506)\ (V_r)$

and:

$L_r = (1.017)\ (D_r)$

Then: $(0.506)\ V_r = (1.017)\ (D_r) = (1.017)\ (0.66372)$

L_r = (1.017) (0.6637) = 0.674 mols/mol feed.

The feed is a boiling point liquid from statement of problem:

q = 1.0

$L_s = L_r + qF$

Basis: 1 mol feed:

$L_{s\ (min)}$ = 0.674 + (1) (1) = 1.674

V_s = 1.332

Determine temperature of rectifying section pinch.

Component	**Dx_{Di}**		
h	0.01072	0.00805	
l	0.428	0.321	
L	0.225	0.169	
Note: 0.225/1.332 = 0.169			

Note: 0.225/1.332 = 0.169	
$K_i = \alpha_i K_B$	
B = reference	

Assume temperature at rectifying pinch. If the components were known, then the overhead dew point and bottoms bubble point could be determined, and from this an approximation could have been made of the pinch temperature. Because these cannot be calculated, trial-and-error must be used to obtain the correct pinch temperature.

Because the 1.017 ≈1.0, continue calculations composition of rectifying pinch:

Component		**Revised Mol Fraction**
2 = h	0.333	0.333/1.017 = 0.328
3 = l	0.580	0.572
4 = L	0.104	0.102
	1.017	1.002

Determine temperature of stripping section pinch:

Component	**Bx_{Bi}**		

		0.10	
1 = H	0.10	1.332 = 0.0752	
2 = h	0.214	0.1605	
3 = l	0.0214	0.0160	

Exact K_B must be between 1.23 and 1.24. Because there is a difference involved in calculation, the result is very sensitive to small changes in K.

Composition of Stripping Pinch:

Component	See Last Col. Above	Revised Mol Fraction
1 = H	0.0794	0.0794/0.9714 = 0.0817
2 = h	0.252	0.2585
3 = 1	0.64	0.659
	0.9714	0.9992

Calculate C_r for each component lighter than light key.

There is only one component lighter than light key in this example, # 4

Reading curve, $C_r = 1.0 = C_n$ (refer to Figure 10-97)

Evaluate C_s for No. 1:

Reading curve, $C_s = 0.875$ (see Figure 10-97)

Now, substitute into Colburn correlation, for check,

$0.684 = (1 - 0.102)\,(1 - 0.0318) = (0.898)\,(0.9682)$

$0.684 \neq 0.868$

Because left side of equation is smaller than right side, assumed was too large. Try a smaller value around 0.95. Right side of equation is not so sensitive to change.

where

L = all components lighter than light key

= sum of all components lighter than light key, does not include light key

= sum of all components heavier than heavy key, does not include heavy key

p = pinch

r = rectifying

s = stripping

l = light key

h = heavy key

H = all components heavier than heavy key, not including the

heavy key

L = all components lighter than light key, not including the light key

C_r, C_s = empirical constants

10.38. Scheibel-Montross Empirical: Adjacent Key Systems: Constant or Variable Volatility [61]

This method has not found as much wide acceptance, in use or in discussion in the literature; nevertheless it allows a direct approximate solution of the average multicomponent system with accuracy of 1–8% on average. If the key components are less than 10% of the feed, the accuracy is probably considerably less than indicated. If a split key system is considered, Scheibel reports fair accuracy when the split components going overhead are estimated and combined with the light key, the balance considered with the heavy key in the L/D relation.

(10-485)

where

x_{lF} = mol fraction of light key in feed

Σx_{FL} = sum of all mol fractions lighter than light key in feed

R' = pseudo minimum reflux

Σx_{FH} = sum of all mol fractions heavier than heavy key in feed

x_{hF} = mol fraction heavy key in feed

α_l = relative volatility of light key to heavy key at feed tray temperature

α_H = relative volatility of components heavier than heavy key at feed tray temperature

α_L = relative volatility of components lighter than light key at feed tray temperatures

x_{it} = mol fraction liquid at intersection of operating lines at minimum reflux. (Calculated or from graph.)
x_{io} = mol fraction light key in overhead expressed as fraction of total keys in overhead

Pseudo minimum reflux:

(10-486)

When the overhead contains only a very small amount of heavy key, the second term in the equation may be neglected.

Intersection of operating lines at Equilibrium Curve:

The proper value for x_{it} is positive and between zero and one. Actually this is fairly straightforward and looks more difficult to handle than is actually the case.

Pseudo ratio of liquid to vapor in feed:

(10-488)

where
x_L = mol fraction of feed as liquid
x_v = mol fraction of feed as vapor
F_L = mols of liquid feed
F_v = mols of vapor feed
ΣF_H = total mols of components heavier than heavy key in feed
ΣF_L = total mol of components lighter than light key in feed.

Example 10-46: Scheibel-Montross Minimum Reflux [61]

A tower has the following all liquid feed composition:

Component	Feed Mols/hr	Overhead Mols/hr	Bottoms Mols/hr
A	30	30.	–
B (light key)	20	19.5	0.5
C (heavy key)	20	0.5	19.5
D	30	–	30.0
	100	50.0	50.0

Relative volatilities referenced to the heavy key, C:

$\alpha_A = 4.0$

$\alpha_B = 2.0 = \alpha_l$

$\alpha_C = 1.0 = \alpha_h$

$\alpha_D = 0.5$

Intersection of Operating Lines:

$x_{it} = 0.610$, or -0.459 (not acceptable)

Pseudo minimum reflux ratio:

R' = 1.472

Minimum reflux ratio:

$(L/D)_{min} = 0.912$

10.39. Minimum Number of Trays: Total Reflux−Constant Volatility

The minimum theoretical trays at total reflux can be determined by the Fenske relation as previously given:

(10-489)

Note that N_{min} is the number of trays in the column and does not include the reboiler. When α varies considerably through out the column, the results will not be accurate using the α_{avg} as algebraic average, and the geometric mean is used in these cases.

For extreme cases it may be necessary to calculate down from the top and up from the bottom until each section shows a fairly uniform temperature gradient between trays. Then the Fenske relation can be used for the remaining trays, using the conditions at the trays calculated as the terminal conditions instead of the actual overhead and bottoms.

10.39.1. Theoretical Number of Trays at Operating Reflux of a Multicomponent Mixture

The method of Gilliland [23] (Figure 10-49a and Figure 10-93) is also used for multicomponent mixtures to determine theoretical trays at a particular operating reflux ratio, or at various ratios. The Brown and Martin [9] curve of Figure 10-98 is also used in approximate the same manner, and produces essentially the same results, but is based on internal vapor and liquid flows.

Figure 10-98

Brown and martin: operating reflux and stages correlated with minimum reflux and stages.

Used and adapted by permission, Van Winkle, M., Oil and Gas Jour. V. 182, Mar. 23 (1953).

The values needed to use the graph include:

(10-490)

where (D/L) = 1/ (L/D)

(10-491)

$L_r = (L/D)\ (D_r)$

(10-492)

(10-493)

Note that when $(L/D)_{min}$ is used as the starting basis, the L_r, L_s, V_r, V_s and their ratios will be for the minimum condition, and correspondingly so when the operating reflux is used.

The combined Fenske-Underwood-Gillilland (FUG) method developed by Frank [100] is shown in Figure 10-95. This relates product purity, actual reflux ratio, and relative volatility (average) for the column to the number of equilibrium stages required. This does not consider tray efficiency. It is perhaps more convenient to use in the design of new columns than for reworking existing columns, and should be used only on adjacent-key systems.

Eduljee [107] evaluated published data and corrected relationships for determining the number of actual trays versus actual reflux with reasonably good agreement:

First attempt:

when

$(S/S_m)\ (R/R_m) = 2.82$

when $R/R_m > 2.0$

(10-494)

If the number of actual trays, S, calculates to be 27 or greater, then

revert to the following for better accuracy:

Second attempt:

when $1.1 < R/R_m \leq 2.0$

(10-495)

when $R/R_m > 2.0$

(10-496)

where:

n = number of theoretical trays in the rectifying section
R = reflux ratio (O/D)
S = number theoretical trays in the column, including reboiler

Subscript

m = minimum

The feed plate location, for either rectifying or stripping sections:

For R/R_m from 1.2 to 3.6:

(10-497)

Hengstebeck [224] presents a technique for locating the feed tray by plotting.

Example 10-47: Operating Reflux Ratio

The minimum reflux ratio $(L/D)_{min}$ has been determined to be 1.017. Using the Brown and Martin graph [9], evaluate the theoretical number of trays at an operating reflux of 1.5 times the minimum. The minimum number of stages was determined to be 22.1 including the reboiler. See Figure 10-98.

The column will have a total condenser. Product rate D is 0.664 mols/mol feed, and the feed is a boiling point liquid.

Minimum values:

$0.506V_r = 1.017\ (D_r) = 1.017\ (0.664)$

$V_r = 1.332$ mols per mol of feed

$L_r = 1.017\ (0.664) = 0.674$ mols/mol feed

$q = 1.0$

$L_s = (0.674) + (1)\ (1) = 1.674$ mols/mol feed

$V_s = 1.332 - (1)\ (1 - 1) = 1.332$ mols/mol feed

Operating values:

Operating $(L/D)_o = (1.5)(1.017) = 1.525$

$L_r = (1.525)\ (0.664) = 1.013$ mols/mol feed

$q = 1.0$

$L_s = 1.013 + (1)\ (1) = 2.013$

$V_s = 1.68 - (1)\ (1 - 1) = 1.68$

$(L/V)_s = 2.013/1.68 = 1.198$

For graph:

Read curve for “greater than 8” minimum equilibrium steps:

at 0.985/1,48 = 0.666

Curve reads: $S_o/S_M = 1.64$

Theoretical stages at reflux (L/D) = 1.525

$S_o = S_M\ (1.64) = 22.1\ (1.64)$

$S_o = 36.2$ stages

Theoretical trays at the operating reflux (L/D) = 1.525

N_o = 36.2 − 1 (for reboi1er) = 35.2 trays in column

10.39.2. Actual Number of Trays

From the theoretical trays at operating reflux the actual trays for installation are determined:

(10-498)

The reboiler is considered 100% efficient, and likewise any partial condenser, if used. Therefore the value N_o represents the theoretical trays or stages in the column proper, excluding the reboiler and partial condenser. E_o represents the overall tray efficiency for the system based upon actual test data of the same or similar systems, or from the plot of Figure 10-54, giving operating information preference (if reliable).

10.39.3. Estimation of Multicomponent Recoveries

Yaws *et al.* [141] present a useful technique for estimating overhead and bottoms recoveries with a very good comparison with tray-to-tray computer calculations. The procedure suggested uses an example from the reference (with permission):

1. Plot relative volatility (α_i) and % desired recovery for LK and HK. Draw a straight line through these two points. The non-key component points will also be on this straight line.
2. Using α_i and the component distribution line, estimate % recovery of non-key components in distillate and bottoms.

From the references [124, 141]:

(10-499)

where

d_i = mols of component i in distillate

b_i = moles of component i in bottoms

α_i = relative volatility of component i

a, b = correlation constants

log (d_i/b_i) vs. log α_i gives a straight line (Figure 10-99). By superimposing a Y_{iD} scale over the d_i/b_i scale, $d_i/b_i = Y_{iD}/Y_{iB}$

Figure 10-99

Estimation of recovery of non-key components using shortcut method of Yaws, et. al., Chem. Eng. Jan. 29, p. 101, (1979).

where

Y_{iD} = % recovery of i in distillate

Y_{iB} = % recovery of i in bottoms

f_i = total mols of component i in distillate and bottoms

Then

(10-500)

(10-501)

From Equation 10-501, Table 10-32 is constructed for selected values of d_i/b_i at various values of Y_{iD} from 99.9% to 0.1%.

10.39.4. Component Recovery Nomograph (Figure 10-99)

A nomograph is constructed by plotting d_i/b_i vs. α_i on log-log graph paper and then superimposing a Y_{iD} scale over the d_i/b_i scale, according to the values given in Table 10-33. The resulting nomograph, relating component recovered and component relative volatility, is given in Figure 10-100. This may be used to estimate component recovery in distillate and bottoms, as follows.

Figure 10-100

Working chart of Yaws, et. al. shortcut method for multicomponent distillation for estimating component recovery in distillate and bottoms.

Used by permission, Yaws et al., Chem. Eng. Jan. 29, p. 101, (1979).

Example 10-48: Estimated Multicomponent Recoveries by Yaws' Method [141] (used with permission)

Component C is to be separated from Component D by distillation. A 95% recovery of both key components (LK, HK) is desired. Saturated-liquid feed composition and relative volatilities (at average column conditions) are given in Table 10-37.

Table 10-37 Material Balance for Estimated Multicomponent Distillation Recoveries for Example 10-48 Using Method of Yaws, Fang, and Patel

Recoveries for Example 10-48 Using Method of Yaws, Fang, and Fuller

Feed component	Composition, mol. fr.	Relative volatility	Keys
A	0.05	3.5	
B	0.20	3.0	
C	0.30	2.3	Light
D	0.25	1.0	Heavy
E	0.15	0.83	
F	0.05	0.65	
Distillate component	**Recovery desired, %**	**Recovery derived, %**	
A		99.72**	
B		99.20**	
C	95	-	
D	5	-	
E		1.30**	
F		0.22*	
Bottoms component	**Recovery desired, %**	**Recovery derived, %**	
A		0.28**	
B		0.80**	
C	5	-	
D	95	-	
E		98.70**	
F		99.78**	
Used by permission, Yaws, C. L., et al., Chem. Eng., Jan. 29 (1979), p. 101, All rights reserved.			
*See calculations.			
**From Figure 10-99.			

Using the graphical shortcut method for component distribution, estimate the recovery of non-key components in distillate and bottoms.

Solution

1. α_i and % desired recovery are plotted for LK and HK (α_c = 2.3, 95% recovery of C in distillate and α_D = 1, 95% recovery of D in bottoms), as shown inFigure 10-99. See Figure 10-100 for working chart. A straight line is then drawn through the two points.
2. Using α_A = 3.5, α_B = 3.0, α_E = 0.83, α_F = 0.65, and the component distribution line, the recovery of non-key components is estimated. The results are shown in Table 10-38.

Table 10-38 Table of Y_{iD} Values for Solving Yaws, Fang, and Patel, Short cut Recoveries Estimate.

Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i
99.9	999	96	24.0	40	0.6670	2	0.02040
99.8	499	94	15.7	30	0.4290	1.0	0.01010
99.6	249	92	11.5	20	0.2500	0.8	0.00806
99.4	166	90	9.00	15	0.1760	0.6	0.00604
99.2	124	85	5.67	10	0.1110	0.4	0.00402
99.0	99.0	80	4.00	8	0.0870	0.2	0.00200
98.5	65.7	70	2.33	6	0.0638	0.1	0.00100
98.0	49.0	60	1.50	4	0.0417		
97.0	32.3	50	1.00	3	0.0309		

Used by permission, Chem. Eng., Yaws, C. L., et al Jan. 29 (1979), p. 101. All rights reserved.

Table 10-39 [141] illustrates the good agreement between the proposed method with the tray-to-tray calculations for Case I-High Recovery: 95% LK recovery in distillate, 94% HK in bottoms; Case

II, Intermediate Recovery: 90% LK recovery in distillate, 85% HK recovery in bottoms; and Case III Low Recovery: 85% LK recovery in distillate, 81 % HK recovery in bottoms.

Table 10-39 Comparison of Yaws, *et al.* Short Cut Nomograph Results vs. Plate-to-Plate Calculations

	Component	Nomograph	Plate to plate	Composition x_i	
				Distillate nomograph	Bottoms plate to plate
Case I High recovery (16 trays)	A	0.0901	0.0901	0.0002	0.0002
	B	0.3588	0.3591	0.0026	0.0023
	C (LK)	0.5197	0.5190	0.0269	0.0278
	D (HK)	0.0271	0.0271	0.5271	0.5271
	E	0.0041	0.0045	0.3314	0.3308
	F	0.0002	0.0002	0.1118	0.1118
Case II Intermediate recovery (13 trays)	A	0.0879	0.0880	0.0016	0.0012
	B	0.3466	0.3464	0.0128	0.0120
	C (LK)	0.4814	0.4770	0.0683	0.0726
	D (HK)	0.0668	0.0682	0.4839	0.4835
	E	0.0155	0.0187	0.3218	0.3187
	F	0.0018	0.0018	0.1116	0.1120
Case III Low recovery (9 trays)	A	0.0866	0.0872	0.0034	0.0028
	B	0.3376	0.3395	0.0250	0.0227
	C (LK)	0.4561	0.4552	0.1015	0.1027
	D (HK)	0.0844	0.0839	0.4606	0.4610
	E	0.0308	0.0295	0.3016	0.3031
	F	0.0045	0.0046	0.1079	0.1077

Used by permission, Chem. Eng., Yaws, C. L., et al Jan. 29 (1979), p. 101, all rights reserved.

Example 10-49: Estimated Multicomponent Recoveries by Yaws' Method [141]

Table 10-40 [141] shows case 1 (High recovery of 11 components), 94.70 % LK recovery and 95.52% HK recovery and case 2 (Low recovery of 13 components), 73.51 % LK recovery and 63.43% HK recovery. Using the shortcut method for component distribution, estimate the recovery of non-key components in distillate and bottoms.

Table 10-40 ▪

Component	f_i	α_i
Case 1. 11 components 94.70 % LK recovery and 95.52 % HK recovery.		
M	0.06	2.5
N	0.03	2.3
O	0.05	2.1
P	0.09	1.75
Q	0.13	1.60
R (LK)	0.15	1.45
S (HK)	0.14	1.00
T	0.11	0.90
U	0.10	0.83
V	0.08	0.75
W	0.06	0.70
Case 2. Low recovery (13 components) 73.51 % LK recovery and 63.43 % HK recovery.		
K	0.01	1.6
L	0.05	1.4
M	0.02	1.3
N	0.12	1.2
O	0.05	1.14

P	0.09	1.12
Q (LK)	0.16	1.10
R (HK)	0.17	1.00
S	0.04	0.98
T	0.11	0.94
U	0.09	0.90
V	0.06	0.80
W	0.03	0.65

Solution

Program PROG 104 is used to determine the recovery of non-key components in distillate and bottoms, using the thermal feed condition of saturated liquid, q = 1. The program (as illustrated earlier) determines the Underwood constant, the minimum reflux ratio, minimum number of stages, actual reflux ratio, the number of theoretical plates and the position of the feed plate. Table 10-41 and Table 10-42 show the results of the program. Comparison of these results with plate-to-plate calculations is shown in Table 10-43. There is good agreement between them in both cases. Further, Microsoft Excel spreadsheet Example 10-49.xls is used to determine the number of theoretical stages using the Erbar-Maddox correlation and the Rusche Equation 10-452.

Table 10-41 Input Data and Computer Results for Multicomponent Fractionation

Data Name: Data104.Dat	
REL	
11	
0.06	2.50
0.03	2.30
0.05	2.10

0.09	1.75
0.13	1.60
0.15	1.45
0.14	1.0
0.11	0.9
0.10	0.83
0.08	0.75
0.06	0.7
0.9470	0.9552
71	1.15

Multicomponent System Fractionation

Component Number	Feed Moles	Rel Volatility Alpha	Distilattle%	Moles	Bottoms %	Moles
1	0.0600	2.5000	99.9991	0.0600	0.0009	0.0000
2	0.0300	2.3000	99.9965	0.0300	0.0035	0.0000
3	0.0500	2.1000	99.9850	0.0500	0.0150	0.0000
4	0.0900	1.7500	99.7243	0.0898	0.2757	0.0002
5	0.1300	1.6000	98.8541	0.1285	1.1459	0.0015
6	0.1500	1.4500	94.7000	0.1421	5.3000	0.0079
7	0.1400	1.0000	4.4800	0.0063	95.5200	0.1337
8	0.1100	0.9000	0.8622	0.0009	99.1378	0.1091
9	0.1000	0.8300	0.2376	0.0002	99.7624	0.0998
10	0.0800	0.7500	0.0471	0.0000	99.9529	0.0800
11	0.0600	0.7000	0.0156	0.0000	99.9844	0.0600

Component Number	Feed Moles	Mol Fract. XF	Distilattle Moles	Mol Fract. XD	Bottoms Moles	Mol. XB
1	0.0600	0.0600	0.0600	0.1182	0.0000	0.0000
2	0.0300	0.0300	0.0300	0.0591	0.0000	0.0000
3	0.0500	0.0500	0.0500	0.0984	0.0000	0.0000
4	0.0900	0.0900	0.0898	0.1767	0.0002	0.0005

5	0.1300	0.1300	0.1285	0.2531	0.0015	0.0030
6	0.1500	0.1500	0.1421	0.2797	0.0079	0.0162
7	0.1400	0.1400	0.0063	0.0124	0.1337	0.2717
8	0.1100	0.1100	0.0009	0.0019	0.1091	0.2216
9	0.1000	0.1000	0.0002	0.0005	0.0998	0.2027
10	0.0800	0.0800	0.0000	0.0001	0.0800	0.1625
11	0.0600	0.0600	0.0000	0.0000	0.0600	0.1219

The feed condition, Q:	1.0
The heavy key compnent number is:	7
Percentage recovery of the light key component	
In the distillate is (%):	94.70
Percentage recovery of the heavy key component	
In the bottoms is (%):	95.52
Total moles in the distillate:	0.5078
Total moles in the bottoms:	0.4922
Total moles in the feed:	1.0000
Underwood constant:	1.1577
Factor for the reflux ratio:	1.1500
Minimum reflux ratio:	2.2977
Actual reflux ratio:	2.6424
Minimum number of	

Minimum number of stages is:	16.0
Number of theoretical plates in the column:	36.2
The position of the feed plate is:	18.9

Table 10-42 Input Data and Computer Results for Multicomponent Fractionation

Data name: Data104.Dat						
REL						
13						
0.01	1.6					
0.05	1.4					
0.02	1.3					
0.12	1.2					
0.05	1.14					
0.09	1.12					
0.16	1.10					
0.17	1.0					
0.04	0.98					
0.11	0.94					
0.09	0.90					
0.06	0.80					
0.3	0.65					
0.7351	0.6443					
81	1.5					
Multicomponent System Fractionation						
		Rel				

Component Number	Feed Moles	Rel Volatility Alpha	Distilattle %	Moles	Bottoms %	Moles
1	0.0100	1.6000	99.9370	0.0100	0.0630	0.0000
2	0.0500	1.4000	99.3979	0.0497	0.6021	0.0003
3	0.0200	1.3000	97.9181	0.0196	2.0819	0.0004
4	0.1200	1.2000	92.3776	0.1109	7.6224	0.0091
5	0.0500	1.1400	83.5591	0.0418	16.4409	0.0082
6	0.0900	1.1200	79.0165	0.0711	20.9835	0.0189
7	0.1600	1.1000	73.5100	0.1176	26.4900	0.0424
8	0.1700	1.0000	35.5700	0.0605	64.4300	0.1095
9	0.0400	0.9800	28.1639	0.0113	71.8361	0.0287
10	0.1100	0.9400	16.2144	0.0178	83.7856	0.0922
11	0.0900	0.9000	8.4782	0.0076	91.5218	0.0284
12	0.0600	0.8000	1.2437	0.0007	98.7563	0.0593
13	0.0300	0.6500	0.0373	0.0000	99.9627	0.0300

Component Number	Feed Moles	Mol Fract. XF	Distilattle Moles	Mol Fract. XD	Bottoms Moles	Mol. XB
1	0.0100	0.0100	0.0100	0.0193	0.0000	0.0000
2	0.0500	0.0500	0.0497	0.0958	0.0003	0.0006
3	0.0200	0.0200	0.0196	0.0378	0.0004	0.0009
4	0.1200	0.1200	0.1109	0.2138	0.0091	0.0190
5	0.0500	0.0500	0.0418	0.0806	0.0082	0.0171
6	0.0900	0.0900	0.0711	0.1371	0.0189	0.0392
7	0.1600	0.1600	0.1176	0.2268	0.0424	0.0880
8	0.1700	0.1700	0.0605	0.1166	0.1095	0.2275
9	0.0400	0.0400	0.0113	0.0217	0.0287	0.0597
10	0.1100	0.1100	0.0178	0.0344	0.0922	0.1914
11	0.0900	0.0900	0.0076	0.0147	0.0284	0.1711
12	0.0600	0.0600	0.0007	0.0014	0.0593	0.1231
13	0.0300	0.0300	0.0000	0.0000	0.0300	0.0623
The feed condition, Q:	1.0					

The feed condition, Q:	1.0
The heavy key component number is :	8
Percentage recovery of the light key component	
In the distillate is (%)	73.51
Percentage recovery of the light key component	
In the bottoms is (%)	64.43
Total moles in the distillate:	0.5186
Total moles in the bottoms:	0.4814
Total moles in the feed:	1.0000
Underwood constant:	1.0438
Factor for the reflux ratio:	1.1500
Minimum reflux ratio:	5.2769
Actual reflux ratio:	6.0684
Minimum number of stages is:	16.9
Number of theoretical plates in the colum:	36.8
The position of the feed plate is:	17.3

Table 10-43 Shortcut versus plate – to – plate calculations.

			% recovery	in dist. $100d_i/f_i$	% recovery	in btms. $100b_i/f_i$	Absolute

Component	f_i	α_i				$100D_i/F_i$	Absolute deviation
			Shortcut	Plate-to-Plate	Shortcut	Plate-to-plate	%
Case 1. High recovery (11 components) 94.70 % LK recovery and 95.52% HK recovery.							
M	0.06	2.5	99.9991	99.999	0.0009	0.001	0.0001
N	0.03	2.3	99.9965	99.996	0.0035	0.004	0.0005
O	0.05	2.1	99.9850	99.986	0.0150	0.014	0.0010
P	0.09	1.75	99.7243	98.698	0.2757	0.302	0.0263
Q	0.13	1.6	98.8541	98.826	1.1459	1.174	0.0281
R (LK)	0.15	1.45	94.7	94.722	5.3000	5.278	0.0220
S (HK)	0.14	1.0	4.48	4.501	95.520	95.499	0.0210
T	0.11	0.9	0.8622	0.83	99.1378	99.170	0.0322
U	0.10	0.83	0.2376	0.221	99.7624	99.779	0.0166
V	0.08	0.75	0.0471	0.042	99.9529	99.958	0.0051
W	0.06	0.7	0.0156	0.013	99.9844	99.987	0.0026
							Avg. 0.0141 %
Case 2. (13 components) 73.51 % LK recovery and 64.43 % HK recovery.							
K	0.01	1.6	99.9253	99.911	0.0747	0.089	0.0143
L	0.05	1.4	99.3285	99.274	0.6715	0.726	0.0545
M	0.02	1.3	97.7574	97.745	2.2426	2.255	0.0124
N	0.12	1.2	92.0941	92.264	7.9059	7.736	0.1699
O	0.05	1.14	83.3346	83.677	16.6654	16.323	0.3424
P	0.09	1.12	78.8802	79.228	21.1198	20.772	0.3478
Q (LK)	0.16	1.10	73.5100	73.896	26.4900	26.104	0.3860
R (HK)	0.17	1.00	36.5700	36.264	63.4300	63.736	0.3060
S	0.04	0.98	29.2394	28.827	70.7606	71.173	0.4124
T	0.11	0.94	17.2097	16.818	82.7903	83.182	0.3917
U	0.09	0.90	9.2141	9.031	90.7859	90.969	0.1831
V	0.06	0.80	1.4349	1.491	98.5651	98.509	0.0561

W	0.03	0.65	0.0474	0.064	99.9526	99.936	0.0166
							Avg. 0.207 %
Note: Case 1: N = 31, N_f = 17, D = 50.8, R = 3.05							
Case 2: N = 28, N_f = 15, D = 52.2, R = 8.5							
(Source: Yaws, C. L., et al., Chem. Eng., p 101, Jan. 299, 1979).							

10.39.5. Shortcut Methods: Reflux and Stages

The shortcut methods allow determination of the number of theoretical plates as a function of reflux ratio, minimum number of plates and minimum reflux. They are used to study the effect of reflux ratio on investment and operating costs with minimum computational effort. These methods are also useful for comparing other variables related to reflux ratio and number of stages. The Underwood [72] and Colburn [12] minimum reflux methods are used for more accurate calculation of minimum reflux, while the Brown-Martin [312] method is a suitable approximation. The Underwood [73] or Fenske methods are useful for estimating minimum number of stages, and the Brown-Martin [312] and Gilliland [23] correlations relate minimum reflux, minimum number of plates, operating reflux and number of theoretical plates.

The minimum number of stages at total reflux as given by Fenske's equation is:

(10-182)

The Fenske's equation has been shown to be rigorous if

(10-502)

where is usually obtained from one of the following approximations:

1.

(10-502a)

2.

(10-502b)

3.

(10-502c)

4.

(10-502d)

5.

(10-502e)

Listed below is the preferred by investigators, which is based upon the above conditions in Equations 10-502a–e.

Method	Recommendation

Method	Recommendation
1	Maddox [112, 266]
2	Van Winkle [74]
3 or 4	King [319]
4	Thompson *et al.* [320], Treybal [321], Wankat [273], Ludwig [271]
4 or 5	Fair [322], Seader and Kurtyka [323], McCormick and Roche [324]

Douglas [325] proposed a criterion for testing the relative volatility approximation as:

(10-503)

When the above inequality is obeyed, the relative volatility is reasonably constant throughout the column, and the simpler approximations such as 2 or 4 are suitable. The Fenske's equation applies to any pair of components and not necessarily to the light key and heavy key.

10.40. Smith–Brinkley (SB) Method [326]

Smith and Brinkley [326] developed a method for estimating the distribution of components in multicomponent separation processes. The method is based on an analytical solution of the finite-difference equations that can be written for multistage separation processes when stages and interstage flow rates are known. The Fenske-Underwood-Gilliland (FUG) method combines Fenske's total-reflux equation and Underwood's minimum reflux equation with a graphical correlation by Gilliland. The latter relates actual column performance to total and minimum reflux conditions for a specified separation between two key components. The method can be used for extraction and absorption processes as well as distillation. The SB and FUG methods are rating and design methods respectively, and both work best when

mixtures are nearly ideal. Only the equations for distillation are presented here, as the derivation of the equations is given by Smith and Brinkley [326] and Smith [65]. For any component i (suffix i omitted in the equation for clarity).

(10-504)

where

b/f = the fractional split of the component between the feed and the bottoms.

N_r = number of equilibrium stages above the feed.

N_s = number of equilibrium stages below the feed.

S_r = stripping factor, rectifying section = .

S_s = stripping factor, stripping section = .

V and L = total molar vapor and liquid flow rates, and the superscript′ denotes the stripping section

G depends on the condition of the feed.

If the feed is mainly liquid:

(10-505)

and the feed stage is added to the stripping section.

If the feed is mainly vapor:

(10-506)

K_i and K'_i are determined at the effective top and bottom section temperatures. If the temperature profile is available (e.g. from a computer simulation), the effective temperature is the arithmetic average of all tray temperatures in the column section. Alternatively, an arithmetic average of the feed-stage and end-stage temperatures is also used.

(10-507)

(10-508)

Application

The Smith-Brinkley method rates a column using the reverse sequence of steps. It takes the number of stages, reflux ratio and actual feed location, and determines the product compositions. Once a base case is given, small changes in variables can readily be made. For example, for a change in feed temperature, half the change is added to t_r and the other half to t_s. For a change in reflux rate, the extra liquid is added to S_s and S_r. If the changes are large, some trial-and-error calculation is required, in which t_s and t_r are varied. This is similar to an operator who adjusts the control temperature by trial and error until the desired end products are achieved. These features make the Smith-Brinkley method valuable for on-line optimization (e.g. using computer or microprocessor control).

Equation 10-505 is for a column with a total condenser. If a partial condenser is used, the number of stages in the rectifying section should be increased by one. The procedure for using the Smith-

Brinkley method is as follows [327]:

1. Estimate the flow rates, L, V and [redacted] from the specified component separations and reflux ratio.
2. Estimate the top and bottom temperatures by calculating the dew and bubble points for assumed top and bottom compositions.
3. Estimate the feed point temperature.
4. Estimate the average component K values in the stripping and rectifying sections.
5. Calculate the values of [redacted] for the rectifying section and [redacted] for the stripping section.
6. Calculate the fractional split of each component, and hence the top and bottom compositions.
7. Compare the calculated with the assumed values and check the overall column material balance.
8. Repeat the calculation until a satisfactory material balance is obtained. The usual procedure is to adjust the feed temperature up and down till a satisfactory balance is obtained.

The SB is a rating method, which is suitable for determining the performance of an existing column rather than a design method, as the number of stages must be known. It can be used for design if estimation of the number of stages is first accomplished by some other method. Equation 10-504 is then used to determine the top and bottom compositions. The estimated stages can thus be adjusted and the computations repeated until the required specifications are achieved. The Gedde-Hengstebeck method (as illustrated earlier) is easier to use and gives satisfactory results for preliminary design in estimating the component splits. Examples of the application of the Smith-Brinkley

method are given by Smith [65].

Commercial process simulators have a wide range of distillation models with varying degree of sophistication. It is essential that the model is chosen for the problem type, the degree of information available and the level of detail required in the solution. The simplest distillation models are shortcut models, which use the Fenske-Underwood-Gilliland (FUG) or Winn-Underwood-Gilliland (WUG) method to determine the minimum reflux and the number of stages. Alternatively they determine the required reflux given a number of stages or the required number of stages for a given reflux ratio. The shortcut methods can also estimate the condenser and reboiler duties and determine the optimum feed stage. The minimum information required to specify a shortcut distillation model is:

- The component recoveries of the light and heavy key components.
- The condenser and reboiler pressures.
- Whether the column has a total or partial condenser.

There are instances where the designers can specify the purities of the light and heavy key components in the distillate and bottoms respectively. However, care must be taken when using purity as a specification, as it is easy to specify infeasible purities or combinations of purities.

The easiest way to use a shortcut distillation model is to start by estimating the minimum reflux ratio and number of stages. The optimum reflux ratio is usually between 1.05–1.25 times the minimum reflux ratio, however, is often used as initial estimate. Once the reflux ratio is specified, the number of stages and optimum feed stage can be determined. The shortcut model results can start as a basis for a rigorous distillation

simulation. Shortcut models can be used to initialize fractionation columns with multiple products, as they are versatile and are solved rapidly. They do not give an accurate prediction of the distribution of non-key components and they do not perform well when there is significant liquid-phase nonideality. The main disadvantage of shortcut models is that they assume constant relative volatility, which is usually calculated at the feed condition. Such models should not be used, if there is significant liquid or vapor phase nonideality, as this assumption does not hold at all well.

10.41. Retrofit design of distillation columns

Since distillation is an energy-intensive process requiring considerable capital investment, retrofit of distillation columns is often carried out in preference to installation of new equipment. Retrofits aim to reuse the existing equipment more effectively in order to increase profit and to reduce cost, for example, by increasing a unit's capacity. When such a revamp is carried out, it is important to make effective utilization of the existing equipment. If the capacity of a systems having two existing columns needs to be increased, then instead of replacing the existing columns with new ones, a third column might be added to the existing columns and the new system reconfigured. In carrying out a retrofit study, retrofit models are employed to fix the existing distillation design, as they are quicker to solve and more robust for optimization, especially when all design variables are being considered simultaneously. They can also be combined with detailed heat-integration models for improving the energy efficiency of distillation systems. Heat integration of distillation columns is reviewed later in the book.

Gadalla *et al.* [328] presented retrofit shortcut models for the design of reboiled and steam stripped distillation columns. These models are particularly valuable for evaluating retrofit design options, and for improving the performance of existing distillation systems. The

models are based upon a modified Underwood method, the Gilliland correlation, the Kirkbride equation, the Fenske equation and the material balances. The models fix both column configurations and operating conditions such as the steam flow rates, and then calculate the product flow rates, temperatures and compositions, and the various heat duties.

These models account for the changes in relative volatility and molar overflow through the column, overcoming the underlying limitations of the previous shortcut models. The models are reliable for very complex configurations, including a – large number of components those having such as crude oil distillation. The models also apply to other distillation applications, such as naphtha fractionation and petrochemical separation. The retrofit models are suited for simple distillation columns, sequences of simple distillation columns, and complex distillation configurations including columns with side-strippers and side-rectifiers. They provide a basis for optimizing and improving operating conditions of existing columns for energy related, economic and environmental benefits. They can be applied to estimate the additional heating and cooling requirements for increased throughput to an existing distillation process. Detailed presentation of these methods is given by Gadalla [329] and Gadalla *et al*. [328].

Example 10-50 Shortcut method using Honeywell UniSim® simulation software

Estimate the number of ideal stages needed in the butane-pentane splitter defined by the compositions given in the table. The column will operate at a pressure of 120 psia (830 kPa), with a reflux ratio of 2.5. The feed is at its boiling point (q = 1).

Component	x_F	**Feed (f_i)**
Propane, C_3	0.05	5
Isobutane, iC_4	0.15	15

Normal butane, nC_4	0.25	25
Isopentane, iC_5	0.20	20
Normal pentane, nC_5	0.35	35

For a specification of not more than 1 mol of the light key in the bottoms product, and not more than 1 mol of the heavy key in the distillate product. Use the Peng-Robinson VLE package.

Solution

The Honeywell UniSim® simulation (Example 10-50) package is used to separate n-butane and i-pentane from the components in the feed entering at its boiling point condition and at 120 psia. The shortcut model in UniSim requires less data than the rigorous model; only five input parameters are required to solve the problem. These are:

1. Two key components: The light key (LK) in the bottoms and the heavy key (HK) in the distillate. The light key is defined as the most volatile component (eg. n-butane) to appear at the bottom of the column and the heavy key is defined as the least volatile component (e.g. i-pentane) to appear at the top of the column. The concentrations of the light and heavy key components must be specified.

   NB: Setting these values too low will result in a large (tall) column with a high reflux, while choosing high values may result in poor product quality and a weak separation.

2. Two pressure specifications: The reboiler and condenser pressures should be specified, as the estimates of condenser and reboiler pressures have a direct relationship on the temperatures that UniSim will estimate in these locations. The reboiler pressure should be greater than the condenser pressure.

   NB: Until the user is satisfied with making estimates about these values an initial guess of the pressures is required, and subsequent checks that the temperature estimates are reasonable

need to be caried out.

The reflux ratio R: The actual reflux ratio should be specified.

Additionally, a fully defined feed stream is required to be setup. Once all the required data are entered, UniSim® simulation software will calculate the remaining column parameters, which can be found on the **Performance tab**. The user should then check the temperature estimates on this page and if necessary adjust the reboiler and condenser pressures.

The following screen shots from UniSim® software illustrate the procedure of carrying out this simulation process.

Step 1. Double click on UniSim icon to display the simulation window as shown in Figure 10-101.

Figure 10-101

Snap shot of UniSim design window.

Step 2. Open the file as a new case as in Figure 10-102 named simulation Basis Manager. Click on the Add button to select the components.

Figure 10-102

Snapshot window of Simulation Basis Manager.

Step 3. Click on the Add Pure button to select the components as highlighted in Figure 10-103. Close the window.

Figure 10-103

Snapshot of the component List View window showing the components.

Step 4. The Simulation Basis Manager window appears as shown in Figure 10-104. Click on Fluid Pkgs tab as shown in Figure 10-104.Then click on the **Add** button.

Figure 10-104

Snapshot of Simulation Basis Manager window.

Step 5. Choose the required property package from the **Property Package Selection** menu. For this example, choose **Peng Robinson** as shown in Figure 10-105. Select **EOSs** from **Property Package Filter** and close the window.

Figure 10-105

Snapshot of Simulation Basis Manager window listing Property Package Selection.

Step 6. Close the window and return to the Simulation Basis Manager window as shown in Figure 10-106. Click on **Enter Simulation Environment** button as shown in Figure 10-106.

Figure 10-106

Snapshot of Simulation Basis Manager window.

Step. 7. Here, the Process Flow Diagram (PFD) window appears, then from the palette, double click on the **blue** arrow and place it on the PFD window. Next, select the shortcut distillation icon in the palette by double clicking on it and place this on the PFD window as shown in Figure 10-107. This is named **T-100**. Double clicking on the arrow -1 gives the worksheet window.

Figure 10-107

Snapshot of the feed stream 1 and distillation column T-100.

Step 8. Enter the **Stream Name**, **Pressure**, **Molar flow** from **Conditions** menu in the Worksheet of the Feed window as shown in Figure10-108.

Figure 10-108

Snapshot of the Feed window showing the **Worksheet** parameters.

Step 9. Next, select Composition menu from the Worksheet in the Feed window. This provides the five components C_3- C_5 that were previously selected for this example.

Step 10. Select **Mole fraction** tab from **Input Composition for Stream: Feed** window and input the data values of the components as shown in Figure 10-110. Click on **OK** button to close the window.

Figure 10-110

Snapshot of **Input Composition For Stream: Feed** window.

Step 11. The total Mole fraction is 1.0 as shown in Figure 10-111. Close this window and double click on the column icon on the PFD window.

Figure 10-111

Snapshot of the Feed window showing the total mole fraction of the components.

Step 12. Double clicking on the Red icon of the column shows the Column window as shown in Figure 10-112. Click on the Design tab and enter the different streams: Feed, Distillate, Bottoms, Reb-Q, Cond-Q. Choose Liquid button for Top Product Phase from Connections as shown in Figure 10-112.

Figure 10-112

Snapshot of **Column** window from **Design** tab in **Connections** menu.

Step 13. Next, select the Parameter menu from Design window. Select light key in Bottoms and its Mole fraction tab, Heavy key in Distillate and its corresponding Mole fraction tab as shown in Figure 10-113. Notice the highlighted red comments displayed **"Unknown Key Components"** at the bottom of the window.

Figure 10-113

Snapshot of **Column** window showing **Parameters** menu.

Step. 14. Input the Parameter values and notice the highlighted red comments displayed **"Unknown Pressure"** as shown in Figure 10-114

Figure 10-114

Further snapshot of **Column** window showing **Parameters** menu.

Step 15. Next, select **Parameters** from the **Design** window. Select Light Key in Component name from the list in Component "**Light key in Bottoms**" (e.g. n-Butane) and enter its value under Mole fraction label. Select Heavy Key component name from the list in Component "**Heavy Key in Distillate**" (e.g. i-Pentane) and enter its corresponding value under Mole fraction label. Enter Condenser and Reboiler Pressure (NB: The value of Reboiler pressure is greater than the value of Condenser pressure). Enter the value of External Reflux Ratio. This gives a highlighted green comment "**OK**" as shown in Figure 10-115.

Figure 10-115

Snapshot of Column window with required Parameters.

Step16. The PFD window shows the results of shortcut simulation with **Feed**, **Distillate** and **Bottoms** and their corresponding values of **Temperature**, **Pressure** and **Molar flow rate** as shown in Figure 10-116.

Figure 10-116

Snapshot of PFD window showing the shortcut method simulation results of Example 10-50.

Step 17. A summary of the printout of the data sheet can be obtained by right clicking on the column icon in Figure 10-117 to display a menu with **Print Datasheet** as shown in Figure 10-117. Then click on the **Print Datasheet** menu.

Figure 10-117

Snapshot of PFD window showing the **Print Datasheet** menu.

Step 18. Click on **Print Datasheet** to display "Select Data block(s) to Print for Shortcut Column" window as shown in Figure 10-118. Click on **Preview** button to view the Shortcut Column datasheet results or alternatively on **Print** button to print

the hard copy of the results.

Figure 10-118

Snapshot of **Select Data block(s) to Print for Shortcut Column**" window.

Figure 10-109.

Snapshot of the Feed window with the selected five components.

Step 19. Display results of the Shortcut Column simulation.

A summary of the results of the shortcut method obtained from both the UniSim simulation software and the developed computer program PROG104 is shown in Table 10-44. The results show that there is good agreement between them both.

Table 10-44 Summary of the results of Example 10-50.

Parameters	Honeywell UniSim®	PROG104
Feed molar flow rate, lbmole/hr	100	100
Distillate molar flow rate, lbmole/hr	44.98	44.99
Bottoms molar flow rate, lbmole/hr	55.02	55.01
Minimum reflux ratio	1.710	1.5360

Minimum reflux ratio	1.719	1.5369
Actual reflux ratio	2.5	2.3054
Minimum number of stages	12.05	13.3
Actual number of stages	21.77	23.6
Optimum feed stage location	10.86	11.8

Figure 10-119 shows the simulation results of Example 10-50.

10.42. Tray-by-Tray for Multicomponent Mixtures

Rigorous tray-by-tray computations for multicomponent mixtures of more than three components can be very tedious, even when made omitting a heat balance. Simulation software packages are type of this adept at dealing with detail and several computational methods are in use.

The direct-solution method of Akers and Wade [1] is among several that attempts to reduce the number of trial-and-error solutions. This has been accomplished and has proven quite versatile in application. The adaptation outlined here modifies the symbols and rearranges some terms for convenient use by the designer [3]. Dew point and bubble point compositions and the plate temperatures can be determined directly. Constant molal overflow is assumed, and relative volatility is held constant over sections of the column.

Rectifying section: reference component is heavy key, x_h

(10-509)

(including x_h)

The compositions of each component are obtained from $(x_i/x_h)_n$ ratio.

The tray temperature is obtained from:

(10-510)

K_h is evaluated at the column pressure by use of suitable K charts.

Stripping section: reference component is heavy key,

(10-511)

The composition of each component on a tray is obtained from $(y_i/y_h)_m$.

The tray temperature is obtained from:

(10-512)

At the column pressure using K charts for the heavy key or reference component.

Procedure

A. Rectifying Section

1. Determine material balance around column, including reflux L, distillate product D, bottoms product B.

a. With total condenser, the reflux composition is equal to the condensed distillate product composition.

b. With a partial condenser, the product D is a vapor, so a dew point must be run on its composition to obtain the liquid reflux composition.

2. Determine top tray temperature for use in relative volatility calculations by running a dew point on the overhead vapor. For total condenser its composition is same as distillate product. For a partial condenser, run a dew point on the column overhead vapor composition as determined by a material balance around the partial condenser, reflux, and product.

3. Determine $(x_i/x_h)_2$, for tray No.2 (second from top), for each component, using the x values for the reflux as the initial x_i (n + 1).

4. Total this column to yield . This equals l/x_h.

5. Determine x_i for each component by:

(10-513)

This is liquid composition on tray.

6. Continue down the column using the composition calculated for the tray above to substitute in Equation 10-509 to obtain (x_i/x_h) for the tray below.

7. Test to determine if α is varying to any great extent by calculating $\alpha_i x_i$ for a test tray. $\Sigma\alpha_i x_i = 1/K_h$. Determine temperature and evaluate corresponding values. Use new α_i if

significantly different.

8. Continue the stepwise calculations until the ratio of light to heavy key on a tray equals (or nearly so) that ratio in the liquid portion of the feed. This is then considered the feed tray.

9. If there are components in the feed and bottoms which do not appear in the overhead product, they must gradually be introduced into the calculations. The estimated position above the feed tray to start introducing these components is determined by:

(10-514)

where

x_{Fi} = mol fraction of a component in the feed that does not appear in the overhead

x_a = small arbitrary mol fraction in the liquid p″ plates above the feed plate

p″ = number of plates above the feed where introduction of components should begin

B. **Stripping Section:**

1. Determine the bubble point temperature of bottoms and composition of vapor, y_{Bi} up from liquid. Calculate relative volatility of light to heavy component at this temperature.

2. From these, calculate the vapor compositions, using Equation 10-511 calculate the ratio (y_i/y_h) for the first tray at the bottom.

3. Total $\Sigma(y_i/y_h)$ to obtain l/y_h

4. Calculate y_i for tray one

(10-515)

$\Sigma y_i = 1.0$

5. Calculate (y_i/y_h) for the next tray, using the y_i values of tray one $(m - 1)$ in the equation to solve for $(y_i/y_h)_m$.
6. Test to determine if α_i is varying significantly by $K_h = \Sigma(y_i/\alpha_i)$. Evaluate the temperature of heavy component at the column bottoms pressure (estimated) using K charts or the equivalent. If necessary, calculate new α_i values for each component at the new temperature. Recheck every two or three trays if indicated.
7. Introduce components lighter than the light key which are not found in the bottoms in the same general manner as discussed for the rectifying section.

(10-516)

where p′ is the number of trays below the feed tray where the component i is introduced in an assumed amount (usually small) x_a. Then x_{Fi} is the mol fraction of the component in the feed.
8. Continue stepwise calculations until ratio of light to heavy keys in the liquid portion of the feed essentially matches the same component ratio in the liquid on one of the trays.
9. The total number of theoretical trays in the column is the sum of those obtained from the rectifying calculations, plus those of the stripping calculations, plus one for the feed tray. This does not include the reboiler or partial condenser as trays in the column.

10.43. Tray-by-Tray Calculation of a Multicomponent Mixture Using a Digital Computer

Multicomponent distillation is much more common in process plants and refineries than the simpler binary systems. Many computer programs have developed to handle the many iterative calculations

required when the system involves three to possibly ten individual components with the necessary degree of accuracy. To solve a multicomponent design properly, there should be both heat and material balance at every theoretical tray throughout the system.

To accommodate the step-by-step recycling and checking for convergence, requires input of vapor pressure relationships (such as those of Wilson, Renon, etc.) using the previously determined constants for latent heat of vaporization (equations) for each component (or enthalpy of liquid and vapor). Specific heat data for each component, and possibly special solubility or Henry's Law deviations may also be required.

There are several valuable references in the literature for developing and applying a multicomponent distillation program, including Holland [26, 27, 169], Prausnitz [52, 53], Wang and Henke [76], Thurston [167], Boston and Sullivan [6], Maddox and Erbar [115], and the pseudo-K method of Maddox and Fling [116]. Convergence of iterative trials to reach a given criterion requires careful evaluation [114]. There are several convergence techniques availiable – some requiring considerably less computational time than others.

Example 10-51: Tray-to-Tray Column Design for Multicomponent Mixture

A column is to be designed to separate the feed given below into an overhead of 99.9 mol % trichloroethylene. The top of the column will operate at 10 psig. Feed temperature is 158°F.

		Overhead		**Bottoms**	
Feed	**Mol Fraction**	**Mols**	**Mol Fraction**	**Mols**	**Mol Fraction**
(A) Trichloroethylene	0.456	0.451	0.999	0.00549	0.010

(B) β Trichloroethane	0.0555	0.00045	0.001	0.05505	0.101
(C) Perchloroethylene	0.3625	…	…	0.36250	0.661
(D) Tetras (1)	0.0625	…	…	0.0625	0.114
(E) Tetras (2)	0.0635	…	…	0.0625	0.114
	1.0000	0.45145	1.000	0.54804	1.000

Note: the material balance for overhead and bottoms is based on:

a. 99.9 mol % trichloroethylene in overhead
b. 1.0 mol% trichloroethylene in bottoms.
c. 1.0 mol feed total
d. Light key = trichloroethylene

Heavy key = β trichloroethane

Determine Overhead Temperature

Because trichloroethylene is 99.9% overhead, use it only to select boiling point from vapor pressure curves at 10 psig overhead pressure = 223°F (1,280 mm Hg abs).

Determine Bottoms Temperature (Bubble Point)

Allowing 10 psi column pressure drop, bottoms pressure = 20 psig (1,800 mm Hg abs)

Component	x_{iB}	**Try t = 320°F Vapor Press. mm Hg**	x_i **(vp.)**	$(yi)_B$
A	0.01	4,500	45	0.0249
B	0.101	2,475	250	0.1382
C	0.661	1,825	1,210	0.67
D	0.114	1,600	183	0.1012

E	0.114	1,050	120	0.0664
			1,808	1.0007
			mm Hg abs.	

This compares quite well with the selected 1,800 mm bottoms pressure. Bottoms temperature is 320°F.

Relative volatilities: Light to Heavy key

At top:

At bottoms:

α (average) = [(2.13) (1.98)]½ = 2.06

Minimum Stages at Total Reflux

Minimum Stages Above Feed

10.44. Thermal Condition of Feed

Feed temperature = 158°F

Calculated bubble point of feed = 266°F at assumed feed tray pressure of 15 psig.

q = 1.298 (Calculations not shown, but handled in a similar manner to the example given in binary section, but all feed components considered, not just keys).

10.45. Minimum Reflux-Underwood Method, Determination of α_{Avg}. For Multicomponent Mixture

Assume pinch temperatures (usually satisfactory because α does not vary greatly) at ⅓ and ⅔ of overall column temperature differences.

Lower pinch = 320 − ⅓ (320 − 223) = 288°F

Upper pinch = 320 − ⅔ (320 − 223) = 255°F

	@ 255°F		@ 288°F		
Component	**v.p.**	**α**	**v.p.**	**α**	**α_i (avg)**
A	2050	2.00	3050	1.91	1.955
B	1025	1.00	1600	1.00	1.00
C	750	0.732	1180	0.737	0.735
D	650	0.634	1035	0.647	0.641
E	390	0.380	650	0.406	0.393

To start, assume θ = 1.113 (it must lie between 1.00 and 1.955).

Component	**x_{Fi}**	**$\alpha_i x_{Fi}$**	**$(\alpha_i - \theta)$**	**$\alpha_i x_{Fi}/(\alpha_i - \theta)$**	**$\alpha_i x_{Fi}/(\alpha_i - \theta)^2$**
A	0.456	0.891	0.842	1.058	1.252
B	0.0555	0.0555	−0.113	−0.491	4.33
C	0.3625	0.266	−0.378	−0.704	1.86
D	0.0625	0.0401	−0.472	−0.085	0.18
E	0.0625	0.0246	−0.720	−0.0342	0.0472
				Σ = −0.2562	Σ 7.669

				Σ = −0.2562	Σ 7.669

$\theta_c = 1.113 - (-0.2562/7.669) = 1.113 + 0.0334$

$\theta_c = 1.146$ (this is sufficiently close to the original, to not require recalculation.)

The correct value of 1.146 should be used.

Check for balance:

$1 - 1.298 = -0.298 = -0.256$

This could be corrected closer if a greater accuracy were needed. It is not as good a match as ordinarily desired.

(for all distillate components)

$= 2.41 + (-0.00685)$

$= 2.404$

$(L/D)_{min} = 2.404 - 1.0 = 1.40$

Operating Reflux and Theoretical Trays-Gilliland Plot

Min trays = $S_M = 12.6$

$(L/D)_{min} = 1.4$

Assume $(L/D)_o$		Read: $(S - S_M)/(S + 1)$	Theoretical stages S
1.4	0	∞	∞
1.6	0.0768	0.546	29
2.0	0.20	0.445	23.5
3.0	0.40	0.312	18.8
4.0	0.52	0.245	17
∞	–	–	12.6

These values are plotted in Figure 10-120. From the curve, the operating $(L/D)_o$ was selected, and the number of corresponding theoretical stages is 19. The spreadsheet Example 10-51.xls calculates the number of theoretical stages using the average value of the volatilities (α_{avg}) of the components at 288°F.

Figure 10-119

Snapshot of shortcut Column simulation results of Example 10-50.

Figure 10-120

Gilliland plot for multicomponent for Example 10-48.

Tray-by-Tray Calculation-Ackers and Wade Method

Rectifying Section, $(L/D)_o = 3:1$

Light key = Trichloroethylene; Heavy key = β Trichloroethane

Relative Volatilities to start: Use average of top and feed

	α_{avg}
A	2.05
B	1.00
C	0.734

Neglect the heavier components than perchloroethylene in the rectifying section.

In order to carry the perchloroethylene, it is assumed at 0.0001 mol fraction in the overhead and reflux, the β-trichloroethane is reduced to 0.0005 mol fraction for these calculations being tighter specifications than the initial calculated balance. The overall effect will be small.

Component	$x_{iD} = x_i$ **(Reflux)**	$(x_i/x_h)_1$	$(x_i)_1$	$(x_i/x_h)_2$	$(x_i)_2$
A	0.9994	975.02	0.9984	545.5	0.9971
B	0.0005	1.00	0.001024	1.0	0.001828
C	0.0001	0.273	0.000280	0.359	0.000656
		Σ = 976.293	0.999704 (sufficiently close)	546.859	

Typical calculations:

For component A: Tray 1

Component B:

Component C:

$(x_A)_1$ = 975.02/976.293 = 0.9984

$(x_B)_1$ = 1.00/976.293 = 0.001024

$(x_c)_1$ = 0.273/976.293 = 0.000280

Tray 2: Component A

Component B:

Component C:

	$(x_i/x_h)_3$	$(x_i)_3$	$(x_i/x_h)_4$	$(x_i)_4$	$(x_i/x_h)_5$	$(x_i)_5$
A	325.24	0.9952	200.81	0.9916	126.61	0.9851
B	1.0	0.00306	1.0	0.004938	1.0	0.007781
C	0.514	0.001573	0.682	0.00337	0.908	0.007065
	326.754		202.492		128.518	

	$(x_i/x_h)_6$	$(x_i)_6$	$(x_i/x_h)_7$	$(x_i)_7$	$(x_i/x_h)_8$	$(x_i)_8$
A	80.60	0.9736	52.05	0.9520	33.97	0.9138
B	1.0	0.01208	1.0	0.01829	1.0	0.0269
C	1.213	0.01465	1.633	0.02987	2.21	0.05945
	82.813		54.683		37.18	

	$(x_i/x_h)_9$	$(x_i)_9$	$(x_i/x_h)_{10}$	$(x_i)_{10}$	$(x_i/x_h)_{11}$	$(x_i)_{11}$
A	22.47	0.8491	15.196	0.7501	7.716	0.5421
B	1.0	0.03779	1.0	0.04936	1.0	0.07026
C	2.994	0.1131	4.061	0.2005	5.516	0.3876
	26.464		20.257		14.232	

Ratio of keys in feed = 0.456/0.0555 = 8.2

Ratio of keys on Tray No. 10 = 0.7501/0.04936 = 15.2

Ratio of keys on Tray No. 11 = 0.5421/0.07026 = 7.7

Tray No. 11 should be used as the feed tray (counting down from the top). Note that since the relative volatility did not change significantly between top to feed trays, the same value was satisfactory over the range.

Stripping Section

Determine V_s: per mol of feed

$L_r = (L/D)\ (D) = 3(0.45145) = 1.35$ mols/mol feed

$L_s = L_r + qF = 1.35 + 1.298\ (1.0) = 2.648$

$V_s = V_r - F(1 - q) = 1.806 - (1.0)\ (1 - 1.298) = 2.104$

$V_s/B = 2.104/0.54804 = 3.84$

Relative volatilities, α_i, determined at average temperature between bottom and feed of the column. Usually the pinch temperature gives equally satisfactory results.

Component	x_{iB}	y_{iB}	$(\alpha_i)_{avg}$	$(y_i/y_h)_1$	$(y_i)_1$	$(y_i/y_h)_2$	$(y_i)_2$
A	0.010	0.0249	1.905	0.319	0.0543	0.552	0.107
B	0.101	0.1382	1.00	1.000	0.170	1.00	0.194
C	0.660	0.6700	0.740	3.800	0.647	3.08	0.597

C	0.660	0.0700	0.740	3.800	0.047	3.08	0.397
D	0.114	0.1012	0.648	0.517	0.088	0.389	0.0754
E	0.114	0.0664	0.411	0.241	0.0411	0.1476	0.0286
Total				5.877		5.1686	

Typical calculations: starting at the bottom and working up the column.

Tray 1: Component A

$(y_i/y_h)_1 = 0.319$

$(y_i)_1 = 0.319/5.877 = 0.0543$

Tray 2: Component A

Continuation of the calculations gives an approximate match of ratio of keys on plate 10 to those in feed. Then feed tray is number 10 from bottom, which is number 11 from top.

Liquid mol fraction ratio from vapor mol fraction ratio:

Ratio on tray no. 9 = 15.018/1.905 = (x_i/x_h) = 7.9

Ratio on tray no. 10 = 19.16/1.905 = 10.05

Ratio in feed = 8.2

Total theoretical trays = 11 + 10 − 1 (common feed tray count)

= 20 not including reboiler

Total theoretical stages = 20 + 1 (reboiler) = 21

This compares with 19 theoretical stages from Gilliland Plot.

Tray Efficiency

Use average column temperature of 271°F and feed analysis.

Component	x_{iF}	**μ, cP**	**μ, x_{iF}, cP**	**vp, mm Hg**	$\alpha_{1/h}$
A	0.456	0.28	0.128	2500	1.94
B	0.0555	0.36	0.020	1290	
C	0.362	0.37	0.134		
D	0.0625	0.40	0.025		
E	0.0625	0.48	0.030		
			Σ = 0.337		
	$\alpha\ \Sigma\ (\mu)\ (x_{iF})$ = 1.94 (0.337) = 0.654				

Using Figure 10-54

Drickamer and Bradford curve, E_o = 46%

O'Connell curve, E_o = 53.8%

In this case, recommend using:

$E_o = (46 + 53.8)/2 = 49.6\%$

Actual trays in column:

$N_{act} = 20/0.496 = 40.3$ trays

From tray-by-tray calculations, feed tray is 10/0.496 = 20.1 trays from bottom, use 20.

Generally, in practice a column would be selected with extra trays, as a contingency *e.g.* a column with a total = 45 trays.

	Number of trays
Rectifying trays	22
Feed	1
Stripping	22
Total	45

Feed nozzles should be located on trays Nos. 21, 23, and 25 up from the bottom tray as No.1.

10.46. Heat Balance-Adjacent Key Systems with Sharp Separations, Constant Molal Overflow

Total Condenser Duty

Refer to Figure 10-121 (System (1)).

1. Assume or set condenser liquid product temperature, t_D.
2. Calculate condensing pressure, with t_D as bubble point (if subcooling exists, and t_D is below bubble point, use bubble point temperature for pressure calculation only).
3. $V_1 = L + D$

(10-517)

(10-518)

(10-519)

(10-520)

4. Calculate t_l and x_1 by dew point on vapor V_1. Then determine H_1, referring to top tray as number one in this case, where

H_1 = total vapor enthalpy above reference datum for sum of all contributing percentages of individual components, i, in stream. Btu/lb, or Btu/mol

h_D = total 1iquid enthalpy above reference datum for sum of all contributing percentages of individual components, i, in product stream. (Also same as reflux), Btu/lb or Btu/mol.

For partial condenser: replace D_{hD} by DH_D in Step 3. A dew point calculation on compositions of y_D (vapor) gives t_D or total pressure. Also get liquid composition x_D (liquid reflux in equilibrium with product vapor y_D). Overhead vapor is the sum of compositions of y_D and x_p. A dew point calculation on this vapor (overhead from tray one top) gives top tray temperature, t_1.

(10-521)

Figure 10-121

Heat balance diagram.

Reboiler Duty

Refer to Figure 10-121 (System (2)).

1. Determine bottoms temperature by bubble point on liquid x_B.
2. From feed condition determine enthalpy.

(10-522)

3. Solve for Q_B, reboiler duty, Btu/hr

(10-523)

where

h_D = total enthalpy of distillate product, Btu/mol or Btu/lb

h_B = total enthalpy of bottoms product, Btu/mol or Btu/lb

h_F = total enthalpy of feed, But/mol or Btu/lb

10.47. Stripping Volatile Organic Chemicals (VOC) from Water

with Air

Li and Hsiao [143] provide a useful approach to the environmental problem of stripping volatile organics from solution in a contaminated water stream by using fresh air as the stripping medium, in order to reduce atmospheric pollution. A number of industrial firms perform this stripping with steam. The mass balance on the VOC component around the column (trayed or packed) is shown in Figure 10-122 uses the symbols of Reference 143.

(10-524)

where

x_o = VOC mol fraction (ratio of number of mols of a specific VOC component in water solution to the total mols of all contaminants contained in the water)

x_N = mol fraction of VOC component in the stripped water

N = number of theoretical trays

y_i = mol fraction VOC component in exiting VOC contaminated air

y_{N+1} = mol fraction VOC component in the incoming fresh air, = 0 for fresh air

L = volumetric flow rate for incoming contaminated water

V = volumetric flow rate for incoming fresh air

V_{min} = minimum fresh air flow required based on slope of operating line L/V on x–y diagram

x_N = mol fraction VOC contaminant in exiting water stream, usually aimed at meeting the environmental regulations

S_{min} = minimum stripping factor at minimum flow rate for air

S_{opt} = optimum stripping factor, where treatment costs are a minimum, referenced to costs of utilities, maintenance, depreciation, labor, as economic conditions change one may need

to adjust S_{opt}, see Reference 143

Figure 10-122

Schematic stripping tower using air to strip organics from water solution.

Adapted and used by permission, Li, K. Y. and Hsiao, K.J., Chem. Eng., V. 98, No. 7, p.114 (1991).

The concentrations of most of the VOC compounds in the contaminated water are usually expressed in ppm, as are the other residual compounds in the water exiting the tower. These are usually small values. As an approximation:

$(L/V_{min}) = K$

where

K = equilibrium constant (varies for each component)

$K = y^*/x^*$

y^* = equilibrium molar fraction of VOC components in air

x^* = equilibrium molar fractions of VOC components in water

Minimum stripping factor at corresponding minimum air flow rate:

(10-525)

$V_{min} = L/K$

The component with the lowest equilibrium constant is the key component in the stripping process, because it yields the largest value of V_{min}. This largest value is the “true” minimum air flow rate, whereas the actual air flow rate should be selected at 1.20 to 2.0 times the minimum. This is based upon an economic balance between the number of actual trays and air flow rate, which determines the column diameter.

It is relevant to evaluate the optimum stripping factor based on related costs, thus:

(10-526)

(10-527)

The Henry's Law constant, H, can be substituted for the equilibrium constant, K, when the system operates at or very close to atmospheric pressure:

(10-528)

where

p^* = the partial pressure, atm, of the contaminant in equilibrium with x^*

Table 10-44 and Table 10-45 provide values for selected Henry's Law constants respectively [143].

Table 10-45 Henry's Law Constants and Optimum Stripping Factors for Selected Organic Compounds for Use with Packed Towers at 25°C (77°F).

Chemicals	Henry's Law constant	x_N/x_o %	L = 30 gpm	L = 300 gpm	L = 3,000 gpm
1,1,2,2 - Tetrachloroethane	24.02	4.75	1.39	1.66	1.84
		0.05	1.88	2.30	2.59
1,1,2-Trichloroethane	47.0	47.5	1.45	1.89	2.32
		0.05	2.00	2.79	3.37
1,2-Dichloroethane	61.2	4.75	1.46	1.97	2.54
		0.05	2.03	2.95	3.73
Propylene dichloride	156.8	4.75	1.6	2.43	3.9
		0.05	2.3	3.9	6.13
Methylene chloride	177.4	4.75	1.57	2.37	3.90
		0.05	2.23	3.87	6.20
Chloroform	188.5	4.75	1.59	2.46	4.10
		0.05	2.28	4.05	6.61
1,1,1-Trichloroethane	273.56	4.75	1.67	2.7	5.08
		0.05	2.43	4.62	8.37
1,2-Dichloroethane	295.8	4.75	1.65	2.68	5.08
		0.05	2.40	4.50	8.40
1,1-Dichloroethane	303.0	4.75	1.67	2.72	5.20
		0.05	2.40	4.63	8.66
Hexachloroethane	547.7	4.75	1.85	3.27	7.74
		0.05	2.7	6.0	13.6
Hexachlorobutadiene	572.7	4.75	1.88	3.48	8.1
		0.05	2.78	6.20	14.27
Trichloroethylene	651.0	4.75	1.82	3.27	7.78
		0.05	2.68	5.87	14.0

1,1-Dichloroethene	834.03	4.75	1.84	3.37	8.50
		0.05	2.70	6.10	15.9
Perchloroethane	1,596.0	4.75	2.10	4.20	13.2
		0.05	3.10	7.90	26.1
Carbon tetrachloride	1,679.17	4.75	2.06	4.2	13.2
		0.05	3.1	7.9	26.45
Used by permission, *Chem. Eng.*, Li, K. Y. and Hsiao, K. J., V. 98, No. 7 (1991), p. 114; all rights reserved.					

The optimum stripping factor, S_{opt}, is expressed as a percent of residue, (100) (x_N/x_o), for water rates of 30 gpm, 300 gpm, and 3,000 gpm.

$$S_{opt} = 1 + {}_aH^b$$

Constants a and b were determined from a linear regression for x_N/x_o = 4.75% and x_N and x_o = 0.05% for the packed and tray towers. The optimum stripping factor decreases as the Henry's Law constant decreases. Due to the complex relationship between cost and performance, the authors [143] recommend caution in attempting to extrapolate from the water flow rate ranges shown.

Example 10-52: Stripping Dissolved Organics from Water in a Packed Tower Using Method of Li and Hsiao [143]

Using a packed tower, remove hexachloroethane (HCE) at a concentration of 110 ppm in water to 0.05 ppm using fresh air operating at essentially atmospheric pressure using a fan/blower generating up 1½ in. water pressure. The concentration of propylene dichloride (PDC) in the contaminated water is 90 ppm, and is to be reduced to 0.05 ppm in the exiting water. The water flow rate = 300 gpm. The required packing (or trays) must be determined by using a vapor-liquid equilibrium plot, setting slope L/V and stepping off the number of stages or transfer units. See Figure 10-122.

From Table 10-45 (Packed Tower):

Hexachloroethane: Henry's Law constant = 547.7 atm

Propylene dichloride: Henry's Law constant = 156.8 atm

1. For hexachloroethane: x_N/x_o = 0.05 ppm/100 ppm = 0.05%
For propylene dichloride: x_N/x_o = 0.05 ppm/100 ppm = 0.05%
2. S_{opt} = 6.0 for HCE, and 3.9 for PDC.
3. For HCE:
V_{min} = L/K = (300) (8.33) (359 scf/mol)/(18 lb/mol) (547.7) = 91.1 scf/minute
For PDC:
V_{min} = L/K = (300 gpm) (8.33) (359)/(18) (156.8) = 317.8 scf/min
4. Use the larger air rate as control required, which is the 317.8 scf/minute required for PDC, to calculate the optimum flow rate.
5. V_{opt} = (3.9) (317.8) = l,239.4 scf/min
$S_{opt} = K/(L/V_{opt})$
$V_{opt} = S_{opt} (V_{min})$
6. Therefore, the operating conditions would be:
L = 300 gpm
V = 1,239 scf/min (minimum, may want to consider actually using 10−15% more for some assurance that the required conditions will be met.
7. Determine the tower diameter based on the flows of (6) above. See Chapter 13, this volume for packed tower design.

10.48. Rigorous Plate–to–Plate Calculation (Sorel Method [311])

The plate to plate calculation by Sorel using mole fractions is outlined as follows:

1. (a) For a total condenser as shown in Figure 10-123, the

temperature of plate 1 is calculated as the dew point temperature of the vapor having the same composition as the distillate D. Therefore, a trial and error calculation is carried out until the relation:

(10-529)

is satisfied. This is carried out by selecting a number of temperatures (usually a minimum of three), and obtaining K or equivalent values at the column pressure and the selected temperature for each component, then summing the y/K values. This is followed by plotting the summation versus temperature and reading the temperature where the summation equals 1.0, as shown in Figure 10-124. The liquid composition in equilibrium with the vapor from plate 1 is calculated by obtaining the K values at the temperature and pressure on plate 1 and by evaluating y/K for each component.

Figure 10-123

Total condenser.

Figure 10-124

Determination of dew point.

(b) For a partial condenser as shown in Figure 10-125, the temperature of the vapor product and reflux liquid must be determined. This temperature is known as the dewpoint of the vapor, and is calculated in the same manner as described in part 1(a). A plot of the summation versus temperature allows the temperature corresponding to the point where the sum equals unity to be read. The composition of the reflux liquid L_o is determined by obtaining K values for each component at the calculated temperature and pressure, and determining .

Figure 10-125

Partial condenser.

2. The quantity of vapor rising from plate 1 is determined by material balance. Since and D are known or fixed,

(10-530)

or

(10-531)

3. The condenser duty or heat removed in the condenser is then calculated by enthalpy balance, since all temperatures and pressures are known.

(10-532)

4. The composition of the vapor V_2 rising to plate 1 (or V_1 in the case of partial condenser) is calculated by material and enthalpy balance. The equations for each case are illustrated as follows:

Total Condenser	Partial Condenser
	(10-533)
	(10-534)
	(10-535)
	(10-536)

With a total condenser and reflux liquid cooled to a temperature

below its bubble point, as shown in Figure 10-126, there are insufficient equations to determine the composition of V_2, quantity of V_2 and L_1, and the temperature T_2. Assuming that the molal latent heat of the mixture is essentially constant, and varies very little with temperature and composition, and that the molal heat capacities of the components in the system are essentially constant, and vary only slightly with temperature, then the following equations can be used to determine the solution:

(10-537)

(10-538)

(10-539)

5. With known V_2, the composition can be determined by Equation 10-534, since the quantities L_1 and D and the compositions and are known.

Note: Enthalpies of the streams may be calculated approximately by the following:

with a total condenser

For a partial condenser:

6. Determine the composition of L_2 in equilibrium with V_2 at T_2 and the column pressure by calculating the equilibrium concentrations by:

(10-540)

7. The vapor V_2 rising to plate 2 can be determined if the partial molal enthalpies of the components in the mixture are known from equations comparable to Equations 10-533 through 10-536. If the Lewis-Matheson assumptions of equimolal overflow and boilup are made, V_3 equals V_2.

8. Determine the composition of V_3 by component and material balance.

9. Calculate the temperature of V_3 by the equation comparable to Equation 10-539.

10. Repeat the calculations plate by plate until a vapor composition is reached which approximates the composition of the vapor portion of the feed, or until a liquid composition is reached that approximates the composition of the liquid portion of the feed.

Note: As a guide, if the calculated ratio of the compositions of the key components approximately matches that of the ratio of keys in the feed, and if the calculated temperature essentially matches the temperature of the feed, the number of equilibrium stages is approximately correct [74].

11. From Figure 10-127, the quantity of liquid and vapor below the feed plate can be determined by:

(10-541)

(10-542)

Figure 10-127

Condition at the feed.

12. Estimate the bubble-point temperature of the bottoms product B, if not determined above.

13. The composition of is the equilibrium composition with B and is calculated by .

14. Since the quantities of and are known, the composition of can be determined by material balance:

(10-543)

15. The temperature of plate n is the bubble-point temperature of L_n and is calculated in the same manner as the reboiler temperature.
16. The calculations are then continued, plate by plate, up the column; determining liquid compositions by material balance, temperature by bubble point computation, and vapor composition by the equilibrium calculations until the liquid and vapor compositions approximate the compositions of the liquid and vapor portions of the feed, or until the key component ratio matches approximately that of the feed. This procedure and observations apply as to the computations proceeding down from the top of the column. The method is sufficient if the key component composition ratio approximately matches that of the feed and the plate temperature is essentially the same as that of the feed.

Figure 10-126

Reflux caused by supercooled liquid reflux from condenser.

10.49. Multiple Feeds and Side Streams for a Binary Mixture

Simple binary distillation columns have been considered with single feed and two products, namely the distillate and bottoms. However, it is possible to have more than one feed or more than two product streams. For example, in the distillation of crude petroleum, several

products such as natural gasoline, liquefied petroleum gas (LPG) propane, butane, motor gasolines, solvent naphtha, aviation fuels, kerosene, diesel oil, gas oils, lube oils, fuel oils and asphalts are all being withdrawn from the column at different points. The McCabeThiele method can be used to analyze such complex configurations.

Two Feed Streams

Consider a typical distillation column as shown in Figure 10-128 having two feeds F_1 and F_2; with an overhead and bottoms product D and B respectively. Figure 10-129 is the related McCabeThiele diagram showing the various operating lines. Each feed has its own q line, *i.e.* feed line. There are three distinct sections in Figure 10-128, namely:

Section I. The top section above the top feed tray.
Section II. The intermediate section between the two feed trays.
Section III. The bottoms section below the lower feed tray.

Figure 10-128

A two-feed column.

Figure 10-129

The McCabeThiele Diagram of a two-feed column.

There are three operating lines in Figure 10 –129, each having different slopes (i.e. L/V ratio).

Side Stream Columns

Some columns receive feed at one point, but have one or more side streams drawing liquid or vapor product over the normal distillate and bottoms products – as in petroleum refineries. Figures 10–130a and c schematically represent a system in which a vapor or liquid side stream may be withdrawn. The removal of liquid from tray N_S reduces the liquid flow rate in the section below the side stream. Hence the L/V ratio is higher above the side stream than below it as shown in Figure 10–130b. Three additional variables, such as flow rate, S, type of side draw (liquid or vapor), and location or composition x_S or y_S must be specified. The operating equation for the middle section can be derived from the mass balances around the top or bottom section of the column. Further, the operating line changes the slope at the side stream tray. The material balance equations at the top, and the side stream assuming constant molal overflow are:

(10-544)

and

(10-545)

Equating the two operating line Equations (10-544) and (10-545), the intersection occurs at:

(10-546)

and since

(10-547)

the point of intersection becomes

The intersection of the lines is:

(10-548)

and intercept occurs at:

(10-549)

Figure 10-130c schematically represents a system where a vapor side stream is withdrawn from tray N_S. The vapor flow rate in the section above N_S is lower than in the section below it. As the liquid flow rate, L remains unchanged, the L/V ratio above the side stream is higher than below it (Figure 10-130aFigure 10-130bFigure 10-130c and Figure 10-130d). For a vapor side stream, the feed line is horizontal (q = 0) at . A balance on vapor flow rates gives:

(10-550)

Figure 10-130a A liquid side stream column.

Figure 10-130b The McCabeThiele diagram with a liquid side stream column.

Figure 10-130c A vapor side stream column.

As L remains unchanged, can be calculated if L and V are known. If a specified value of x_S or y_S is desired, the problem becomes one of a trial and error.

Example 10-53

An ethanol–water mixture is to be separated by continuous fractionation in a sieve-plate distillation column operated at 101.325 kPa. A feed flow rate of 1000 kmol/h containing 28 mole % ethanol is to be separated into two ethanol-rich streams; a distillate containing 82.5 mole % ethanol, and a side stream containing 60 mol % ethanol. It is required to recover 98% of the ethanol in the feed in these two

product streams.

The molar flow rate of the distillate will be twice that of the side stream. The feed will be liquid at its boiling point, and the product streams will be removed as liquids, also at their respective boiling points. An external reflux ratio of 2.5 to 1.0 will be applied at the top of the column. Stating clearly any assumptions, calculate:

a. The composition of the bottoms product.

b. Estimate the number of sieve plates required if the plate efficiency is 60%.

c. Indicate briefly how a distillate containing 99.5 mole % ethanol could be obtained.

Solution

Assuming constant molal overflow:

(a) overall material

(1)

The component balance on the more volatile component at the top of the column and side stream is:

(2)

Since it is required to recover 98 % of the ethanol in the feed, then:

Since the molar flow rate of the distillate is twice that of the side stream:

The feed is at its boiling point, q = 1 or

274.4 = 0.825 D + 0.3D

Hence,

D = 243.9 kmol/h

and

S = 243.9/2 = 121.95 kmol/h

Material balance on the upper operating line:

Component balance is:

The equation of the upper operating line is:

where R = L/D or V = (R + 1)D

The intercept of the upper operating line (UOL) is

The bottoms flow rate is:

= 1000 – 243.9 – 121.95

= 634.15 kmol/h

But 2 % of the ethanol in the feed is removed at the bottoms, that is:

= 0.0088

(b) The intermediate slope is assumed a constant molal overflow.

Mass balance at the top of the column is:

= 609.75 + 243.9

= 853.65 kmol/h

Side stream is liquid at its boiling point; therefore a mass balance on the side stream is:

But

Therefore

= 609.75 – 121.95

= 487.8 kmol/h

The slope of the side stream is:

The number of sieve plates from the McCabeThiele diagram (Figure 10-131) is 13. The actual number of sieve plates with a plate efficiency of 60% is:

(c) Use an azeotrope extracting agent *e.g.* benzene, cyclohexane, or adjust the pressure to change the composition of the azeotrope. This will require a greater ΔP shift.

Figure 10-130d The McCabeThiele diagram with a vapor side stream column.

Figure 10-131

McCabeThiele diagram for Example 10-53.

Example 10-54

The feed to a continuous distillation column operating at atmospheric pressure contains 75 mole per cent pyridine, 17 mole per cent α-picoline and 8 mole per cent β-picoline. The distillate contains 99 mole per cent pyridine, 1 mole per cent α-picoline, whilst the residue 20 mole per cent pyridine, 53.67 mole per cent α-picoline and 26.33 mole per cent β-picoline. The relative volatility of pyridine with respect to β-picoline is 2.32 and the relative volatility of α-picoline with respect to β-picoline is 1.54 at the mean column temperature. If the minimum reflux and actual operating reflux are 2.65:1 and 5:1 respectively, calculate:

a. the approximate number of theoretical plates to accomplish the above separation

b. the approximate location of the feed plate.

Solution

a. The approximate number of theoretical plates to accomplish the above separation is as follows:

Using Fenske's equation to determine the minimum number of stages at total reflux:

(10-182)

where:

The minimum number of stages is:

= 13.6 (14 stages)

R_m = 2.65, R = 5.10

Using Gilliland correlation, the abscissa is:

(10-219)

The corresponding ordinate on Gilliland correlation is:

(10-220)

$N - N_m = y\ (N + 1)$

(10-453)

The number of theoretical plates is

N = 19 stages.

b. The approximate location of the feed plate is:

Using the Kirkbride's equation for the feed plate location:

(10-462)

where

m = number of theoretical stages above the feed plate.

p = number of theoretical stages below the feed plate.

Basis of the feed flow rate, F = 100 kmol/h.

The overall mass balance of the column is:

(1)

(2)

(3)

Component balance on the more volatile component, pyridine (MVC) is:

(4)

(5)

Substituting Eq. (3) into Eq. (5) gives:

(6)

(7)

B = 30.37 kmol/h

D = 100 – 30.37

= 69.63 kmol/h

Substituting these values in Kirkbride's feed plate location equation gives:

(8)

(9)

(10)

(11)

Substituting Eq. 10 into Eq. 11 gives:

(12)

3.133p = 19

p = 6.06 stages

m = 12.94 (13 stages)

The feed plate is between 12th – 13th stage from the top of the

column.

Using Akashah *et al.* [317] Equation 10-462b to determine the location of the feed plate from Equation 10-462 or 10-462a:

(10-462)

or

(10-462a)

(10-462b)

where N = 19 and p = 6.06 from Kirkbride's Equation 10-462 or 10-462a, we have:

= 12.29 (12 stages)

(10-462c)

p = 19 – 12

= 7 stages

10.50. Chou and Yaws Method [96]

This method for multicomponent distillation involving more than one

feed and more than one side stream requires a reliable minimum reflux ratio.

In summary, the calculation procedure is as presented by the authors.

For the systems rated above, the minimum reflux ratio is [96]:

(10-551)

(10-552)

This includes recognizing the contribution from the feed (R_F), “other feeds” (R_{OF}), and sidestreams (R_S). The R_F portion is determined by assuming no other feeds or sidestreams are present. The R_{OF} and R_S parts represent the summation of the contributions of other feeds and sidestreams to the overall column minimum reflux ratio. The calculation sequence consists basically of three steps, here reproduced by permission of Chemical Engineering, Chou and Yaws, April 25, 1988, All rights reserved [96]:

1. Determine Underwood θ for each feed, using equation involving feed concentration:

(10-553)

2. Determine the “minimum reflux ratio” candidate for each feed:

(10-554)

where:

3. Compare the candidates for minimum reflux ratio. The candidate having the largest (maximum) numerical value is the minimum reflux ratio for the column.

Example 10-55: Distillation with Two Sidestreams Feeds

Data for Example 10-55, which includes two sidestreams.

		Mole Fraction				
		Feed		**Distillate**	**Sidestream**	
Component	**Relative Volatility α**	**z_i , F_1**	**z_i , F_2**	**x_D**	**z_i , S_1**	**z_i , S_2**
1 (LK)	2.0	0.38	0.45	0.999	0.975	0.900
2 (HK)	1.0	0.20	0.30	0.001	0.025	0.099
3 (HK + 1)	0.5	0.22	0.25	0	0	0.001
Feeds: F_1 = 50 mol/hr, q_{F1} = 1 (saturated liquid)						
F_2 = 100 mol/hr, q_{F2} = 0 (saturated vapor)						
Sidestreams: S_1 = 20 mol/hr, q_{S1} = 1 (saturated liquid)						
S_2 = 20 mol/hr, q_{S2} = 1 (saturated liquid)						

Distillate: D = 36 mol/hr

Minimum reflux and other results for Example 10-55.

Column representation of results of Example 10-55.

1.	UNDERWOOD THETAS:	
	FOR FEED 1	THETA (1) = 1.164
	FOR FEED 2	THETA (2) = 1.485
2.	MINIMUM REFLUX CANDIDATES:	
	FOR FEED 1	RMIN (1) = 3.450271
	FOR FEED 2	RMIN (2) = 4.375502
3.	TRUE MINIMUM REFLUX RATIO:	
	RMIN = 4.38	

where

B = bottoms flowrate, mol/h

c = number of components

D = distillate flowrate, mol/h

F = flowrate of feed, mol/h

F_j = flowrate of feed j, mol/h

F_{FR} = factor for contribution of other feed flow to minimum reflux

F_{FRj} = factor for contribution of feed j flow to minimum reflux

F_{SR} = factor for contribution of side stream flow to minimum

reflux,

$_{FSR,k}$ = factor for contribution of sidestream k flow to minimum reflux

HK = heavy key component

L = liquid flowrate, mol/h

LK = light key component

nf = number of feeds

ns = number of sidestreams

m = number of sidestreams above feed n

q_F = thermal condition of feed

q_S = thermal condition of sidestream

R = reflux ratio

R_F = feed component of minimum reflux

$R_{F,n}$ = feed component of minimum reflux for feed n

R_{OF} = other feeds component of minimum reflux

R_{min} = minimum reflux ratio

R_S = sidestream component of minimum reflux

S = flowrate of sidestream, mol/h

S_k = flowrate of sidestream k, mol/h

V = vapor flowrate, mol/h

x_i = mole fraction of component i in liquid

y_i = mole fraction of component i in vapor

$z_{i,F}$ = mole fraction of component i in feed

$z_{i,Fj}$ = mole fraction of component i in feed j

$z_{i,S}$ = mole fraction of component i in sidestream

$z_{i,Sk}$ = mole fraction of component i in sidestream k

α = relative volatility

θ = Underwood parameter

Subscripts

B = bottoms

D = distillate
F = feed
F_j = feed j
F_n = intermediate feed

10.51. Optimum Reflux Ratio and Optimum Number of Trays Calculations

Van Winkle and Todd have developed shortcut graphical methods which are very useful for estimating the optimum reflux ratio RR_{opt}, and optimum number of stages N_{opt}, such to minimize operating and investment costs for a given separation of multicomponent systems in a fractionation operation. Their methods were based on:

1. Using the shortcut methods of Fenske for the minimum number of stages as shown in Figure 10-132.

Figure 10-132

Graphical representation of Fenske equation for estimating minimum plates
(Source: P. Chattopadhyay [293]).

2. Using the Underwood method for the minimum reflux ratio (Figure 10-133Figure 10-134Figure 10-135 and Figure 10-136).

Figure 10-133

Underwood's parameter (θ) as a function of the ratio of the keys in the

feed
(Source: P. Chattopadhyay [293]).

Figure 10-134

Dependence of on θ for the light key and lighter components for calculating underwood's (RR_m + 1) (Source: P. Chattopadhyay [293]).

Figure 10-135

Dependence of on θ for the light key and lighter components for calculating underwood's (RR_m + 1) (Source: P. Chattopadhyay [293]).

Figure 10-136

Dependence of on θ for the light key and lighter components for calculating underwood's (RR_m + 1)

(Source: P. Chattopadhyay [293]).

3. Using Gilliland's correlations for determining the relationship between theoretical stages and reflux ratio to their minimum value.

This method has been tested on several hundred cases of multicomponent distillation in which α, , percent recovery of LK in the distillate and percent recovery of the HK in the bottoms were varied over a practical range. The following observations were made about the result:

Effect of Pressure: Pressure exerts no appreciable effect on the optimum values of total number of trays and reflux ratio.

Effect of Feed Rate: The optimum reflux ratio RR_{opt} and optimum number of stages N_{opt} vary with the changes in feed flow rates. The value of known as the tray factor increases with the increase in feed rate while the value of known as the reflux factor decreases at the same time.

Material of Construction: The materials of construction mainly used are plain carbon steel and stainless steel; Monel metal of tower and trays has a direct effect on and . However, with the increase in the cost of material M_c, the optimum values of and

decrease.

Utility Cost: The inclusion of utility cost (e.g. cost of steam + cost of cooling water + cost of electricity) in reducing the total cost greatly affects the RR_{opt} and N_{opt}. When the utility cost (also known as the operating cost) is taken into account, optimum level of always increases whereas the optimum level of always decreases.

Correlations

The overall procedure is based on the correlation of and for a given reference case. A base case is selected and then correction factors are correlated to take into account the effects of feed rates, cost of material and the number of years of equipment amortization.

Originally, the selected base case is:

A 3m diameter column of plain carbon steel as the material of construction, an operating pressure of 696 kPa and 2 year amortization. The optimum value of for both the reference case and the range of conditions was correlated as a function of the Fenske separation factor, using α as the parameter as illustrated in Figure 10-137 to estimate values.

Figure 10-137

Correlation of the ratio of optimum to minimum number of trays as a function of Fenske separation factor
(Source: P. Chattopadhyay [293]).

To account for the effects of various feed rates, the feed correction factor ϕ was correlated as a function of feed rate, using α as the parameter as shown in Figure 10-138. Figure 10-138 is used to evaluate the correction factor ϕ for and ϕ is used to calculate as:

(10-555)

Γ is the correction factor for for the depreciation period and material of construction. This correction factor is obtained from Figure 10-139 which plots Γ as a function of for various values of α. is obtained from Figure 10-140 and correction factor for feed ϕ is obtained from Figure 10-141; correction factor for amortization Δ is determined from Figure 10-142.

Figure 10-138

Feed rate correction factor for $(N/N_{mim})_{ref}$

(Source: P. Chattopadhyay [293]).

Figure 10-139

Correction factor for $(N/N_{mim})_{ref}$ for the number of years amortized (n), divided by the cost of materials (M_c)

(Source: P. Chattopadhyay [293]).

Figure 10-140

Correlation of optimum is to minimum reflux ratio for the reference case $(RR/RR_{mim})_{ref}$

(Source: P. Chattopadhyay [293]).

Figure 10-141

Feed rate correction factor for $(RR/RR_{mim})_{ref}$ (Source: P. Chattopadhyay

[293]).

Figure 10-142

Correction factor for $(RR/RR_{mim})_{ref}$ for the depreciation period and material of construction
(Source: P. Chattopadhyay [293]).

Procedure

In order to determine the optimum combination of reflux ratio and the number of stages based on minimum yearly operating cost, the following parameters are required:

1. Feed rate.
2. Feed condition and feed properties.
3. Operating pressure.
4. Operating temperatures.
5. Relative volatilities of the components.
6. Specified separation of key components.
7. Material costs.
8. Amortization years.

The algorithm is as follows:

1. Determine the value of N_{min}. Use either Figure 10-132 or Fenske equation (10-182).
2. Determine RR_{min} from Underwood's method. Use Figure 10-133 to determine the Underwood's parameter θ. Alternatively use computer program PROG 104 or the spreadsheet program

Example 10-41.xls. Use Equations 10-437 and 10-438 or Figure 10-134Figure 10-135 and Figure 10-136 to determine RR_{min}.

3. Obtain the reference value from Figure 10-137, using Fenske separation factor for the correct α.

4. Determine the feed correction factor ϕ from Figure 10-138. Use the correct feed rate (F) and α.

5. The amortization time (n) and the material cost factor (M_c) are used to obtain the correction factor (Γ) for depreciation and material of construction, from Figure 10-139. Calculate the optimum value of from:

(10-555)

6. Using the value of N_{min} from step 1 and the value of the RHS of Equation 10-555 in step 5, determine the optimum value of the number of theoretical stages required.

7. From Figure 10-140 determine on the basis of empirical separation factor (plotted in the abscissa).

8. From Figure 10-141 determine correction factor φ for the feed. Use the correct feed rate (F) and α.

9. For the selected amortization time (n) and material cost factor (M_c), read out the correction factor Δ for from Figure 10-142. Calculate the optimum value of

from

(10-556)

10. With RR_{min} from step 2 and the value of the RHS of Equation 10-556, the optimum value of reflux ratio RR_{opt} is determined.

Generally, the optimum reflux ratio for a multicomponent system is obtained by comparing operating (energy) costs with the capital costs and then optimizing the two. Results of previous economic designs have been compiled, compared and reduced to a relationship of or equivalently based on the type of heat sink employed. Table 10-46 shows the economic ranges of reflux to minimum reflux and the resulting values of stages to minimum stages.

Table 10-46 Economic and vs. Heat Sink Source

Method of condensing the reflux		
Low-level refrigeration (−300 to 150°F)	1.05–1.10	2.0–3.0
High-level refrigeration	1.10–1.20	1.8–2.0

(−150 to 50°F)		
Cooling water at cost of circulation and limited treating	1.20–1.5	1.0–1.8
Air cooling	1.40–1.5	1.6–1.8

Example 10-56: Minimum and Optimum values of theoretical trays and reflux ratio

A multicomponent system comprising of four hydrocarbons: isobutane (iC_4H_{10}), n-butane (nC_4H_{10}), isopentane (iC_5H_{12}), and n-pentane (nC_5H_{12}) is to be separated as specified in the table.

Component	**Feed moles**	**Equilibrium constant K**	**Relative volatility α**	**Distillate moles**	**Bottoms moles**
iC_4H_{10}	0.06	2.15	2.5749	0.0596	0.0004
nC_4H_{10} (LK)	0.17	1.70	2.0359	0.1615	0.0085
iC_5H_{12} (HK)	0.32	0.835	1.0000	0.0160	0.3040
nC_5H_{12}	0.45	0.70	0.8383	0.0054	0.4446
Total	1.00			0.2425	0.7575

The column operating pressure = 103 kPa

Top temperature = 420 K

Bottom temperature = 470 K

95 per cent recovery of both keys is desired.

Feed is saturated liquid at its boiling point (q = 1).

Take: n-butane as light key component

i-pentane as heavy key component

Input Data

Material of construction	= Carbon steel for all equipment
Material cost factor	= 1
Depreciation period	= 10 years
Feed rate	= 500 kmol/h
Tray efficiency	= 85%

Determine the following:

.

Solution

Table 10-47 gives the computer results of PROG 104.

Table 10-47 Computer Output for Multicomponent System Fractionation.

Data name: Data104.Dat	
EQUIL	
0.06	2.15
0.17	1.70
0.32	0.835
0.45	0.70

0.45	0.70
0.95	0.95
3.1	1.3

Component number	Feed moles	Rel. volatility alpha	Distillate %	Moles	Bottoms %	Moles
1	0.0600	2.5749	99.2532	0.0596	0.7468	0.0004
2	0.1700	2.0359	95.0000	0.1615	5.0000	0.0085
3	0.3200	1.0000	5.0000	0.0160	95.0000	0.3046
4	0.4500	0.8383	1.2067	0.0054	98.7633	0.4446

Component number	Feed moles	Mol fract. XF	Distillate moles	Mol fract. XD	Bottoms moles	Mol. fract. XB
1	0.0600	0.0600	0.0596	0.2456	0.0004	0.0006
2	0.1700	0.1700	0.1615	0.6660	0.0085	0.0112
3	0.3200	0.3200	0.0160	0.0660	0.3040	0.4013
4	0.4500	0.4500	0.0054	0.0224	0.4446	0.5869

The feed condition, Q:	1.0
The heavy key component number is:	3
Percentage recovery of the light key component in the distillate is (%):	95.00
Percentage recovery of the heavy key component in the bottoms is (%):	95.00
Total moles in the distillate:	0.2425
Total moles in the bottoms:	0.7575
Total moles in the feed:	1.0000
Underwood constant:	1.6213
Factor for the reflux ratio:	1.3000

Minimum reflux ratio:	2.8036
Actual reflux ratio:	3.6447
Minimum number of stages is:	8.3
Number of theoretical plates in the column:	16.2
The position of the feed plate is:	6.7

From Table 10-47:

Minimum reflux ratio, RR_{min} = 2.8036

Minimum number of stages, N_{min} = 8.3

Underwood constant,

Relative volatility,

The following procedure is used to calculate N_{opt} and R_{opt} :

1. Graphical representation of Fenske equation for estimating minimum plates as:

from Figure 10-137:

2. Underwood constant, θ = 1.6213

3. Correlation of the ratio of optimum to minimum number of

trays as a function of Fenske separation factor from Figure 10-137

4. Correction factor ϕ for feed rate using Figure 10-138

F = 500 kmol/h

log (F) = log (500) = 2.698

From Figure 10-138, ϕ = 0.97

5. Correction factor Γ for for the number of years amortized (n), divided by the cost of materials (M_c).

n = 10 years, M_c = 1.0

,

From Figure 10-139, Γ = 1.185

6. Optimum value of is:

= (0.97) (1.185) (2.75)

= 3.16

7. Optimum number of theoretical stages, where N_{min} = 8.3 (Table 10-47).

= (8.3) (3.16)

= 26.24

8. Correlation of optimum to minimum reflux ratio for the reference case using composition values of LK and HK in Table 10-47.

= 2.25058

From Figure 10-140,

9. Correction factor φ for feed rate from Figure 10-141,

log (F) = 2.70 at

$\varphi = 1.04$

10. Depreciation and material of construction correction factor.

n = 10 years, M_c = 1.0, n/Mc = 10,

From Figure 10-142, $\Delta = 0.9375$

= 1.0335

11. Optimum reflux ratio, RR_{opt}, where RR_{min} = 2.8036 (Table 10-47).

= (2.8036) (1.0335)

= 2.897

Generally, the reflux ratio affects the cost of the column, both in the number of trays and the diameter. It also affects the cost of operation, *i.e.* the costs of heat and cooling supply and power for the reflux pump. Therefore, the basis for choice of an optimum reflux ratio involves an economic balance. Table 10-48 summarizes an economic balance of optimum reflux ratio for typical petroleum fraction distillation. The conclusion may be drawn that the optimum reflux ratio is about 1.2 times the minimum, and also the number of trays is about twice the minimum. Figure 10-143 shows the temperature profiles of iC4, nC4, iC5, nC5 through the column using Sorel method

of plate to plate calculation.

Table 10-48 Economic Optimum Reflux Ratio For Typical Petroleum Fraction Distillation near 1 atm.[a]

Source: Happen and Jordan, Chemical Process Economics, Decker, New York, 1975.

	Factor for optimum reflux										
	$f = (R_{opt}/R_m) - 1$								**Factor for optimum trays**		
	$R_{opt} = (1 + f) R_m$								**N_{opt}/N_m**		
	$N_m = 10$			**$N_m = 20$**			**$N_m = 50$**		**N_m = 10**	**N_m = 20**	**N_m = 50**
		Rm			**R_m**		**R_m**		**R_m**	**R_m**	**R_m**
	1	**3**	**10**	**1**	**3**	**10**	**1**	**10**	**1 to 10**	**1 to 10**	**1 to 10**
Base case	0.20	0.12	0.10	0.24	0.17	0.16	0.31	0.21	2.4	2.3	2.1
Payout time 1 yr.	0.24	0.14	0.12	0.28	0.20	0.17	0.37	0.24	2.2	2.1	2.0
Payout time 5 yr.	0.13	0.09	0.07	0.17	0.13	0.10	0.22	0.15	2.7	2.5	2.2
Steam cost $0.30/M lb	0.22	0.13	0.11	0.27	0.16	0.14	0.35	0.22	2.3	2.1	2.0
Steam cost $0.75/M lb	0.18	0.11	0.09	0.21	0.13	0.11	0.29	0.19	2.5	2.3	2.1
G_a = 50 lb mole/(hr)(sq.ft)	0.06	0.04	0.03	0.08	0.06	0.05	0.13	0.08	3.1	2.8	2.4

[a]The "base case" is for payout time of 2 yr, steam cost of $0.05/1000 lb, vapor flow rate Ga = 15 lb mol/(hr)(sq. ft.). Although the capital and utility costs are prior to 1975 and are individually far out of date, the relative costs are roughly the same so the conclusions of this analysis are not far out of line. Conclusion: For systems with nearly ideal VLE, R is approx. $1.2R_{min}$ and N is approx. 2.0

N_{min}.

Figure 10-143

Typical concentration profiles in separation of light hydrocarbon mixtures when no substantial inversions of relative volatilities occur
(Source: Van Winkle, Distillation, McGraw-hill, New York, 1967).

10.52. Tower Sizing for Valve Trays

Many types of trays are used in both fractionating and absorption columns. In a fractionating column, bubble caps with weirs and downcomers maintain a liquid level on the trays. Liquid flows across the tray via the downcomer and then across the next tray in the opposite direction, whilst vapor flows up through the cap and the slots, thus mixing with the liquid. Figure 10-144 and Figure 10-145 show a section of a column distributor and different types of packings. Figure 10-146 illustrates vapor flow through the bubble cap, sieve and valve trays. The riser in a bubble cap can help to prevent liquid from “weeping” through the vapor passage. Sieve or valve trays control weeping by vapor velocity. The bubble cap tray has the highest turndown ratio designed to be 8:1 to 10:1. Such trays are commonly used in glycol dehydrating columns [290].

Figure 10-144

Nutter Engineering MTS - 109 multipan two stage liquid distributor.

Courtesy of Nutter Engineering.

Figure 10-145

Nutter ring random packing.

Courtesy of Nutter Engineering.

Figure 10-146

Flow through vapor passages. (A) Vapor flow through bubble cap. (B) Vapor flow through perforations. (C). Vapor flow through valves.

(Source: Van Winkle, Distillation, McGraw-Hill, New York, 1967).

Valve and sieve trays are favored because of their lower cost and increased capacity compared to bubble cap trays for a given tower diameter. Sieve or perforated trays are plates with holes for vapor passage. Their simple construction makes them the least expensive of the three designs. Generally, sieve tray shave a higher capacity, but at low vapor rates, they are susceptible to “weeping” or dumping of the liquid through the holes. Furthermore, sieve trays are preferred in fouling applications, or when solids are present. They can have large holes which are highly resistant to clogging, and are easy to clean. In

addition, its turndown ratio is limited.

Trayed columns give satisfactory operation over a wide range of liquid and vapor loadings. Correspondingly, valve trays can maintain high efficiencies over a wider range of operating liquid and gas rates than sieve trays. Valve units are more mechanically complex than sieve trays, and therefore are more expensive to make. The liquid and vapor rates can vary independently over a broad range in the column resulting in satisfactory operation. At low vapor rates, unsatisfactory tray dynamics may be evidenced by dumping of liquid, uneven distribution and vapor pulsation. Alternatively, at high vapor rates, the tower floods as the liquid backs up in the downcomers. Poor vapor-liquid contact can result at low liquid rates. Correspondingly, at high liquid rates, flooding and dumping can also result as liquid capacities are exceeded in the downcomers. Valve trays are designed to have better turndown properties than sieve trays, and therefore are more flexible for use with varying feed rates. At the design vapor rate, valve trays have about the same efficiency as sieve trays. They can be designed for a lower pressure drop (ΔP) than sieve trays. However, they are susceptible to fouling or plugging, if dirty solutions are distilled.

In these examples, valve trays are used for sizing a column. The procedure uses the Tower sizing graphs, and tables from Nutter [291] and Blackwell's [292] correlation.

The Equations

The following expressions are used for the design of valve trays columns:

The density radical R_D is defined by:

(10-557)

The velocity at zero liquid load is:

(10-558)

where

(10-559)

A, B, C, and D are constants

A = 0.22982

B = 0.44605

C = 0.03452

D = 0.00415

E = 0.00017

Equation 10-558 is valid for .

The tray spacing factor is defined by:

(10-560)

where Equation (10-560) is valid for

. The constants G and H for different tray spacing are given in Table 10-49. Figure 10-147 illustrates the plots of versus for different tray spacing.

Table 10-49 Values of G and H for each tray spacing.

Tray spacing (inches)	**G**
12	0.77174
18	0.93655
24	for all values of X
30	0.98057
36	0.96583
	when

Figure 10-147

Plots of tray spacing vs. ratio of surface tension to vapor density.

By permission of Nutter Engineering.

The operating volumetric vapor flow rate Q_V is:

(10-561)

The operating volumetric liquid flow rate Q_L is:

(10-562)

The bubbling area A_b is:

(10-563)

Note: If , then replace 0.5 for in Equation (10-563). Table 10-50 gives recommended values of foam factor and residence time.

Table 10-50 Recommended Values for Foam Factor and Residence Time.

Fractionators	Foam factor	Downcomer res. time, sec.
Straight run		
Hydrocarbon	1.0	4.0
Low MW alcohols	1.0	3.5
Crude fractionator	1.0	4.5
Special		
Rich oil (top) demeth. or deeth.	0.85	4.5

Rich oil (top) demeth. or deeth.	0.85	4.5
Rich oil (BTM) demeth. or deeth.	1.0	4.5
Hydrocarbon still (top)	1.0	4.0
Hydrocarbon still (BTM)	1.0	4.5
MEA-DEA still	0.85	4.5
Glycol-DGA still	0.8	4.5
Sulfinol still	1.0	5.0
stripper	0.9	4.0
Sour water stripper	0.5–0.7	4.5
stripper	1.0	3.0
Refrigerated demeth. or deeth. (Top)	0.8	8.0
Refrigerated demeth. or deeth. (BTM)	1.0	5.0
Absorbers		
Oil – ambient temp. (above)	0.85	4.5
Oil – low temp. (below)	0.95	4.5
DGA-DEA-MEA (contactor)	0.75	4.5
Glycol contactor	0.65	5.0
Sulfinol contactor	1.0	5.0
Vacuum towers		
Crude vacuum	1.0	5.0
By permission of Nutter Engineering.		

Using Table 10-50 to obtain both the foam factor (Foam fac) and the downcomer residence time τ, the downcomer area A_{DC} is:

(10-564)

The expression for the safety factor is:

(10-565)

Equation (10-565) is valid for . where

K = 0.91146
M = −0.03821

The tower area A_T is:

(10-566)

Table 10-51 shows the tower diameter versus its cross sectional area, and Table 10-52Table 10-53Table 10-54 and Table 10-55 give the recommended downcomer area and bubbling area distribution ratios.

Table 10-51 Tower Diameter vs. CrossSectional Areas. By Permission of Nutter Engineer.

Tower Diameter		**Area**
Feet	**Inches**	**Sq. Ft.**
2′-0″	24″	3.142
2′-6″	30″	4.909

3′-6″	36″	7.069
3′-6″	42″	9.621
4′-0″	48″	12.566
4′-6″	54″	15.904
5′-0″	60″	19.635
5′-6″	66″	23.758
6′-0″	72″	28.274
6′-6″	78″	33.183
7′-0″	84″	38.485
7′-6″	90″	44.179
8′-0″	96″	50.266
8′-6″	102″	56.745
9′-0″	108″	63.617
9′-6″	114″	70.882
10′-0″	120″	78.540
10′-6″	126″	86.590
11′-0″	132″	95.033
11′-6″	138″	103.869
12′-0″	144″	113.098
12′-6″	150″	122.719
13′-0″	156″	132.733
13′-6″	162″	143.139
14′-0″	168″	153.938
14′-6″	174″	165.130
15′-0″	180″	176.715
15′-6″	186″	188.692
16′-0″	192″	201.062
16′-6″	198″	213.825
17′-0″	204″	226.980
17′-6″	210″	240.528
18′-0″	216″	254.469

18′-6″	222″	268.803
19′-0″	228″	283.529
19′-6″	234″	298.648
20′-0″	240″	314.159
20′-6″	246″	330.064
21′-0″	252″	346.361
21′-6″	258″	363.050
22′-0″	264″	380.133
22′-6″	270″	397.608
23′-0″	276″	415.476
23′-6″	282″	433.736
24′-0″	288″	452.389
24′-6″	294″	471.435
25′-0″	300″	490.874
25′-6″	306″	510.705
26′-0″	312″	530.929
26′-6″	318″	551.546
27′-0″	324″	572.555
27′-6″	330″	593.957
28′-0″	336″	615.752
28′-6″	342″	637.940
29′-0″	348″	660.520
29′-6″	354″	683.493
30′-0″	360″	706.858
30′-6″	366″	730.617
31′-0″	372″	754.768
31′-6″	378″	779.311
32′-0″	384″	804.248
32′-6″	390″	829.577
33′-0″	396″	855.299
33′-6″	402″	881.413

34′-0″	408″	907.920
34′-6″	414″	934.820
35′-0″	420″	962.113
35′-6″	426″	989.798
36′-0″	432″	1017.876
36′-6″	438″	1046.347
37′-0″	444″	1075.210
37′-6″	450″	1104.466
38′-0″	456″	1134.115
38′-6″	462″	1164.156
39′-0″	468″	1194.591
39′-6″	476″	1225.417

Table 10-52
Downcomer Dimensions.
By Permission of Nutter Engineering.

H/DIA	**L/DIA**	**A_d/A_t**
.0000	.0000	.0000
.0005	.0447	.0000
.0010	.0632	.0001
.0015	.0774	.0001
.0020	.0894	.0002
.0025	.0999	.0002
.0030	.1094	.0003
.0035	.1181	.0004
.0040	.1262	.0004
.0045	.1339	.0005
.0050	.1411	.0006
.0055	.1479	.0007
.0060	.1545	.0008

.0060	.1543	.0008
.0065	.1607	.0009
.0070	.1667	.0010
.0075	.1726	.0011
.0080	.1782	.0012
.0085	.1836	.0013
.0090	.1889	.0014
.0095	.1940	.0016
.0100	.1990	.0017
.0105	.2039	.0018
.0110	.2086	.0020
.0115	.2132	.0021
.0120	.2178	.0022
.0125	.2222	.0024
.0130	.2265	.0025
.0135	.2308	.0027
.0140	.2350	.0028
.0145	.2391	.0030
.0150	.2431	.0031
.0155	.2471	.0033
.0160	.2510	.0034
.0165	.2548	.0036
.0170	.2585	.0037
.0175	.2622	.0039
.0180	.2659	.0041
.0185	.2695	.0042
.0190	.2730	.0044
.0195	.2765	.0046
.0200	.2800	.0048
.0205	.2834	.0050
.0210	.2868	.0051
.0215	.2901	.0053

.0220	.2934	.0055
.0225	.2966	.0057
.0230	.2998	.0059
.0235	.3030	.0061
.0240	.3061	.0063
.0245	.3092	.0065
.0250	.3122	.0067
.0255	.3153	.0069
.0260	.3183	.0071
.0265	.3212	.0073
.0270	.3242	.0075
.0275	.3271	.0077
.0280	.3299	.0079
.0285	.3328	.0081
.0290	.3356	.0083
.0295	.3384	.0085
.0300	.3412	.0087
.0305	.3439	.0090
.0310	.3466	.0092
.0315	.3493	.0094
.0320	.3520	.0096
.0325	.3546	.0098
.0330	.3573	.0101
.0335	.3599	.0103
.0340	.3625	.0105
.0345	.3650	.0108
.0350	.3676	.0110
.0355	.3701	.0112
.0360	.3726	.0115
.0365	.3751	.0117
.0370	.3775	.0119

.0375	.3800	.0122
.0380	.3824	.0124
.0385	.3848	.0127
.0390	.3872	.0129
.0395	.3896	.0132
.0400	.3919	.0134
.0405	.3943	.0137
.0410	.3966	.0139
.0415	.3989	.0142
.0420	.4012	.0144
.0425	.4035	.0147
.0430	.4057	.0149
.0435	.4080	.0152
.0440	.4102	.0155
.0445	.4124	.0157
.0450	.4146	.0160
.0455	.4168	.0162
.0460	.4190	.0165
.0465	.4211	.0168
.0470	.4233	.0171
.0475	.4254	.0173
.0480	.4275	.0176
.0485	.4296	.0179
.0490	.4317	.0181
.0495	.4338	.0184
.0500	.4359	.0187
.0505	.4379	.0190
.0510	.4400	.0193
.0515	.4420	.0195
.0520	.4441	.0198
.0525	.4461	.0201

.0530	.4481	.0204
.0535	.4501	.0207
.0540	.4520	.0210
.0545	.4540	.0212
.0550	.4560	.0215
.0555	.4579	.0218
.0560	.4598	.0221
.0565	.4618	.0224
.0570	.4637	.0227
.0575	.4656	.0230
.0580	.4675	.0233
.0585	.4694	.0236
.0590	.4712	.0230
.0595	.4731	.0242
.0600	.4750	.0245
.0605	.4768	.0248
.0610	.4787	.0251
.0615	.4805	.0254
.0620	.4823	.0257
.0625	.4841	.0260
.0630	.4859	.0263
.0635	.4877	.0266
.0640	.4895	.0270
.0645	.4913	.0273
.0650	.4931	.0276
.0655	.4948	.0279
.0660	.4966	.0282
.0665	.4983	.0285
.0670	.5000	.0288
.0675	.5018	.0292
.0680	.5035	.0295

.0685	.5052	.0298
.0690	.5069	.0301
.0695	.5086	.0304
.0700	.5103	.0308
.0705	.5120	.0311
.0710	.5136	.0314
.0715	.5153	.0318
.0720	.5170	:0321
.0725	.5186	.0324
.0730	.5203	.0327
.0735	.5219	.0331
.0740	.5235	.0334
.0745	.5252	.0337
.0750	.5268	.0341
.0755	.5284	.0344
.0760	.5300	.0347
.0765	.5316	.0351
.0770	.5332	.0354
.0775	.5348	.0358
.0780	.5363	.0361
.0785	.5379	.0364
.0790	.5395	.0368
.0795	.5410	.0371
.0800	.5426	.0375
.0805	.5441	.0378
.0810	.5457	.0382
.0815	.5472	.0385
.0820	.5487	.0389
.0825	.5502	.0392
.0830	.5518	.0396
.0835	.5533	.0399

.0840	.5548	.0403
.0845	.5563	.0406
.0850	.5578	.0410
.0855	.5592	.0413
.0860	.5607	.0417
.0865	.5622	.0421
.0870	.5637	.0424
.0875	.5651	.0428 .
.0880	.5666	.0431
.0885	.5680	.0435
.0890	.5695	.0439
.0895	.5709	.0442
.0900	.5724	.0446
.0905	.5738	.0449
.0910	.5752	.0453
.0915	.5766	.0457
.0920	.5781	.0460
.0925	.5795	.0464
.0930	.5809	.0468
.0935	.5823	.0472
.0940	.5837	.0475
.0945	.5850	.0479
.0950	.5864	.0483
.0955	.5878	.0486
.0960	.5892	.0490
.0965	.5906	.0494
.0970	.5919	.0498
.0975	.5933	.0501
.0980	.5946	.0505
.0985	.5960	.0509
.0990	.5973	.0513

.0995	.5987	.0517

Table 10-53
Downcomer Dimensions
By Permission of Nutter Engineering.

H/DIA	L/DIA	A_d/A_t
.1000	.6000	.0520
.1005	.6013	.0524
.1010	.6027	.0528
.1015	.6040	.0532
.1020	.6053	.0536
.1025	.6066	.0540
.1030	.6079	.0544
.1035	.6092	.0547
.1040	.6105	.0551
.1045	.6118	.0555
.1050	.6131	.0559
.1055	.6144	.0563
.1060	.6157	.0567
.1065	.6170	.0571
.1070	.6182	.0575
.1075	.6195	.0579
.1080	.6208	.0583
.1085	.6220	.0587
.1090	.6233	.0591
.1095	.6245	.0595
.1100	.6258	.0598
.1105	.6270	.0602
.1110	.6283	.0606

.1115	.6295	.0610
.1120	.6307	.0614
.1125	.6320	.0619
.1130	.6332	.0623
.1135	.6344	.0627
.1140	.6356	.0631
.1145	.6368	.0635
.1150	.6380	.0639
.1155	.6392	.0643
.1160	.6404	.0647
.1165	.6416	.0651
.1170	.6428	.0655
.1175	.6440	.0659
.1180	.6452	.0663
.1185	.6464	.0667
.1190	.6476	.0671
.1195	.6488	.0676
.1200	.6499	.0680
.1205	.6511	.0684
.1210	.6523	.0688
.1215	.6534	.0692
.1220	.6546	.0696
.1225	.6557	.0701
.1230	.6569	.0705
.1235	.6580	.0709
.1240	.6592	.0713
.1245	.6603	.0717
.1250	.6614	.0721
.1255	.6626	.0726
.1260	.6637	.0730
.1265	.6648	.0734

.1270	.6659	.0738
.1275	.6671	.0743
.1280	.6682	.0747
.1285	.6693	.0751
.1290	.6704	.0755
.1295	.6715	.0760
.1300	.6726	.0764
.1305	.6737	.0768
.1310	.6748	.0773
.1315	.6759	.0777
.1320	.6770	.0781
.1325	.6781	.0785
.1330	.6791	.0790
.1335	.6802	.0794
.1340	.6813	.0798
.1345	.6824	.0803
.1350	.6834	.0807
.1355	.6845	.0811
.1360	.6856	.0816
.1365	.6866	.0820
.1370	.6877	.0825
.1375	.6887	.0829
.1380	.6898	.0833
.1385	.6908	.0838
.1390	.6919	.0842
.1395	.6929	.0847
.1400	.6940	.0851
.1405	.6950	.0855
.1410	.6960	.0860
.1415	.6971	.0864
.1420	.6981	.0869

.1425	.6991	.0873
.1430	.7001	.0878
.1435	.7012	.0882
.1440	.7022	.0886
.1445	.7032	.0891
.1450	.7042	.0895
.1455	.7052	.0900
.1460	.7062	.0904
.1465	.7072	.0909
.1470	.7082	.0913
.1475	.7092	.0918
.1480	.7102	.0922
.1485	.7112	.0927
.1490	.7122	.0932
.1495	.7132	.0936
.1500	.7141	.0941
.1505	.7151	.0945
.1510	.7161	.0950
.1515	.7171	.0954
.1520	.7180	.0959
.1525	.7190	.9063
.1530	.7200	.0968
.1535	.7209	.0973
.1540	.7219	.0977
.1545	.7229	.0982
.1550	.7238	.0986
.1555	.7248	.0991
.1560	.7257	.0996
.1565	.7267	.1000
.1570	.7276	.1005
.1575	.7285	.1009

.1580	.7295	.1014
.1585	.7304	.1019
.1590	.7314	.1023
.1595	.7323	.1028
.1600	.7332	.1033
.1605	.7341	.1037
.1610	.7351	.1042
.1615	.7360	.1047
.1620	.7369	.1051
.1625	.7378	.1056
.1630	.7387	.1061
.1635	.7396	.1066
.1640	.7406	.1070
.1645	.7415	.1075
.1650	.7424	.1080
.1655	.7433	.1084
.1660	.7442	.1089
.1665	.7451	.1094
.1670	.7460	.1099
.1675	.7468	.1103
.1680	.7477	.1108
.1685	.7486	.1113
.1690	.7495	.1118
.1695	.7504	.1122
.1700	.7513	.1127
.1705	.7521	.1132
.1710	.7530	.1137
.1715	.7539	.1142
.1720	.7548	.1146
.1725	.7556	.1151
.1730	.7565	.1156

.1735	.7574	.1161
.1740	.7582	.1166
.1745	.7591	.1171
.1750	.7599	.1175
.1755	.7608	.1180
.1760	.7616	.1185
.1765	.7625	.1190
.1770	.7633	.1195
.1775	.7642	.1200
.1780	.7650	.1204
.1785	.7659	.1209
.1790	.7667	.1214
.1795	.7675	.1219
.1800	.7684	.1224
.1805	.7692	.1229
.1810	.7700	.1234
.1815	.7709	.1239
.1820	.7717	.1244
.1825	.7725	.1249
.1830	.7733	.1253
.1835	.7742	.1258
.1840	.7750	.1263
.1845	.7758	.1268
.1850	.7766	.1273
.1855	.7774	.1278
.1860	.7782	.1283
.1865	.7790	.1288
.1870	.7798	.1293
.1875	.7806	.1298
.1880	.7814	.1303
.1885	.7822	.1308

.1890	.7830	.1313
.1895	.7838	.1318
.1900	.7846	.1323
.1905	.7854	.1328
.1910	.7862	.1333
.1915	.7870	.1338
.1920	.7877	.1343
.1925	.7885	.1348
.1930	.7893	.1353
.1935	.7901	.1358
.1940	.7909	.1363
.1945	.7916	.1368
.1950	.7924	.1373
.1955	.7932	.1378
.1960	.7939	.1383
.1965	.7947	.1388
.1970	.7955	.1393
.1975	.7962	.1398
.1980	.7970	.1403
.1985	.7977	.1409
.1990	.7985	.1414
.1995	.7992	.1419
.2000	.8000	.1424
.2005	.8007	.1429
.2010	.8015	.1434
.2015	.8022	.1439
.2020	.8030	.1444
.2025	.8037	.1449
.2030	.8045	.1454
.2035	.8052	.1460
.2040	.8059	.1465
.2045	.8067	.1470

.2045	.8067	.1470
.2050	.8074	.1475
.2055	.8081	.1480
.2060	.8089	.1485
.2065	.8096	.1490
.2070	.8103	.1496
.2075	.8110	.1501
.2080	.8118	.1506
.2085	.8125	.1511
.2090	.8132	.1516
.2095	.8139	.1521
.2100	.8146	.1527
.2105	.8153	.1532
.2110	.8160	.1537
.2115	.8167	.1542
.2120	.8174	.1547
.2125	.8182	.1553
.2130	.8189	.1558
.2135	.8196	1563
.2140	.8203	.1568
.2145	.8210	.1573
.2150	.8216	1579
.2155	.8223	.1584
.2160	.8230	.1589
.2165	.8237	.1594
.2170	.8244	.1600
.2175	.8251	.1605
.2180	.8258	.1610
.2185	.8265	.1615
.2190	.8271	.1621
.2195	.8278	.1626
.2200	.8285	.1631

.2200	.8285	.1631
.2205	.8292	.1636
.2210	.8298	.1642
.2215	.8305	.1647
.2220	.8312	.1652
.2225	.8319	.1658
.2230	.8325	.1663
.2235	.8332	.1668
.2240	.8338	.1674
.2245	.8345	.1679
.2250	.8352	.1684
.2255	.8358	.1689
.2260	.8365	.1695
.2265	.8371	.1700
.2270	.8378	.1705
.2275	.8384	.1711
.2280	.8391	.1716
.2285	.8397	.1721
.2290	.8404	.1727
.2295	.8410	.1732
.2300	.8417	.1738
.2305	.8423	.1743
.2310	.8429	.1748
.2315	.8436	.1754
.2320	.8442	.1759
.2325	.8449	.1764
.2330	.8455	.1770
.2335	.8461	.1775
.2340	.8467	.1781
.2345	.8474	.1786
.2350	.8480	.1791
.2355	.8486	.1797

.2355	.8486	.1797
.2360	.8492	.1802
.2365	.8499	.1808
.2370	.8505	.1813

Table 10-54
Downcomer Dimensions.
By Permission of Nutter Engineering.

H/DIA	L/DIA	A_d/A_t
.2375	.8511	.1818
.2380	.8517	.1824
.2385	.8523	.1829
.2390	.8529	.1835
.2395	.8536	.1840
.2400	.8542	1845
.2405	.8548	.1851
.2410	.8554	.1856
.2415	.8560	.1862
.2420	.8566	.1867
.2425	.8572	.1873
.2430	.8578	.1878
.2435	.8584	.1884
.2440	.8590	.1889
.1445	.8596	.1895
.2450	.8602	.1900
.2455	.8608	.1906
.2460	.8614	.1911
.2465	.8619	.1917
.2470	.8625	.1922
.2475	.8631	.1927

.2480	.8637	.1933
.2485	.8643	.1938
.2490	.8649	.1944
.2495	.8654	.1949
.2500	.8660	.1955
.2505	.8666	.1961
.2510	.8672	.1966
.2515	.8678	.1972
.2520	.8683	.1977
.2525	.8689	.1983
.2530	.8695	.1988
.2535	.8700	.1994
.2540	.8706	.1999
.2545	.8712	.2005
.2550	.8717	.2010
.2555	.8723	.2016
.2560	.8728	.2021
.2565	.8734	.2027
.2575	.8745	.2038
.2580	.8751	.2044
.2585	.8756	.2049
.2590	.8762	.2055
.2595	.8767	.2060
.2600	.8773	.2066
.2605	.8778	.2072
.2610	.8784	.2077
.2615	.8789	.2083
.2620	.8794	.2088
.2625	.8800	.2094
.2630	.8805	.2100
.2635	.8811	.2105

.2640	.8816	.2111
.2645	.8821	.2116
.2650	.8827	.2122
.2655	.8832	.2128
.2660	.8837	.2133
.2665	.8843	.2139
.2670	.8848	.2145
.2675	.8853	.2150
.2680	.8858	.2156
.2685	.8864	.2161
.2690	.8869	.2167
.2695	.8874	.2173
.2700	.8879	.2178
.2705	.8884	.2184
.2710	.8890	.2190
.2715	.8895	.2195
.2720	.8900	.2201
.2725	.8905	.2207
.2730	.8910	.2212
.2735	.8915	.2218
.2740	.8920	.2224
.2745	.8925	.2229
.2750	.8930	.2235
.2755	.8935	.2241
.2760	.8940	.2246
.2765	.8945	.2252
.2770	.8950	.2258
.2775	.8955	.2264
.2780	.8960	.2269
.2785	.8965	.2275
.2790	.8970	.2281

.2795	.8975	.2286
.2800	.8980	.2292
.2805	.8985	.2298
.2810	.8990	.2304
.2815	.8995	.2309
.2820	.8999	.2315
.2825	.9004	.2321
.2830	.9009	.2326
.2835	.9014	.2332
.2840	.9019	.2338
.2850	.9028	.2349
.2855	.9033	.2355
.2860	.9038	.2361
.2865	.9043	.2367
.2870	.9047	.2872
.2875	.9052	.2378
.2880	.9057	.2384
.2885	.9061	.2390
.2890	.9066	.2395
.2895	.9071	.2401
.2900	.9075	.2407
.2905	.9080	.2413
.2910	.9084	.2419
.2915	.9089	.2424
.2920	.9094	.2430
.2925	.9098	.2436
.2930	.9103	.2442
.2935	.9107	.2448
.2940	.9112	.2453
.2945	.9116	.2459
.2950	.9121	.2465

.2955	.9125	.2471
.2960	.9130	.2477
.2965	.9134	.2482
.2970	.9139	.2488
.2975	.9143	.2494
.2980	.9148	.2500
.2985	.9152	.2506
.2990	.9156	.2511
.2995	.9161	.2517
.3000	.9165	.2523
.3005	.9170	.2529
.3010	.9174	.2535
.3015	.9178	.2541
.3020	.9138	.2547
.3025	.9187	.2552
.3030	.9191	.2558
.3035	.9195	.2564
.3040	.9200	.2570
.3045	.9204	.2576
.3050	.9208	.2582
.3055	.9212	.2588
.3060	.9217	.2593
.3065	.9221	.2599
.3070	.9225	.2605
.3075	.9229	.2611
.3080	.9233	.2617
.3085	.9237	.2623
.3090	.9242	.2629
.3095	.9246	.2635
.3100	.9250	.2640
.3105	.9254	.2646

.3110	.9258	.2652
.3115	.9262	.2658
.3125	.9270	.2670
.3130	.9274	.2676
.3135	.9278	.2682
.3140	.9282	.2688
.3145	.9286	.2693
.3150	.9290	.2699
.3155	.9294	.2705
.3160	.9298	.2711
.3165	.9302	.2717
.3170	.9306	.2723
.3175	.9310	.2729
.3180	.9314	.2735
.3185	.9318	.2741
.3190	.9322	.2747
.3195	.9326	.2753
.3200	.9330	.2759
.3205	.9333	.2765
.3210	.9337	.2771
.3215	.9341	.2777
.3220	.9345	.2782
.3225	.9349	.2788
.3230	.9352	.2794
.3235	.9356	.2800
.3240	.9360	.2806
.3245	.9364	.2812
.3250	.9367	.2818
.3255	.9371	.2824
.3260	.9375	.2830
.3265	.9379	.2511

.3270	.9382	.2842
.3275	.9386	.2848
.3280	.9390	.2854
.3285	.9393	.2860
.3290	.9397	.2866
.3295	.9401	.2872
.3300	.9404	.2878
.3305	.9408	.2884
.3310	.9411	.2890
.3315	.9415	.2896
.3320	.9419	.2902
.3325	.9422	.2908
.3330	.9426	.2914
.3335	.9429	.2920
.3340	.9433	.2926
.3345	.9436	.2932
.3350	.9440	.2938
.3335	.9443	.2944
.3360	.9447	.2950
.3365	.9450	.2956
.3370	.9454	.2962
.3375	.9457	.2968
.3380	.9461	.2974
.3385	.9464	.2980
.3390	.9467	.2986
.3400	.9474	.2998
.3405	.9478	.3004
.3410	.9481	.3010
.3415	.9484	.3016
.3420	.9488	.3022
.3425	.9491	.3028

.3430	.9494	.3034
.3435	.9498	.3040
.3440	.9501	.3046
.3445	.9504	.3053
.3450	.9507	.3059
.3455	.9511	.3065
.3460	.9514	.3071
.3465	.9517	.3077
.3470	.9520	.3083
.3475	.9524	.3089
.3480	.9527	.3095
.3485	.9530	.3101
.3490	.9533	.3107
.3495	.9536	.3113
.3119	.3500	.9539
.3505	.9543	.3125
.3510	.9546	.3131
.3515	.9549	.3137
.3520	.9552	.3143
.3525	.9555	.3150
.3530	.9558	.3156
.3535	.9561	.3162
.3540	.9564	.3168
.3545	.9567	.3174
.3550	.9570	.3180
.3555	.9573	.3186
.3560	.9576	.3192
.3565	.9579	.3198
.3570	.9582	.3204
.3575	.9585	.3211
.3580	.9588	.3217

.3585	.9591	.3223
.3590	.9594	.3229
.3595	.9597	.3235
.3600	.9600	.3241
.3605	.9603	.3247
.3610	.9606	.3253
.3615	.9609	.3259
.3620	.9612	.3265
.3625	.9614	.3272
.3630	.9617	.3278
.3635	.9620	.3284
.3640	.9623	.3290
.3645	.9626	.3296
.3650	.9629,	.3302
.3655	.9631	.3308
.3660	.9634	.3315
.3665	.9637	.3321
.3675	.9642	.3333
.3680	.9645	.3339
.3685	.9648	.3345
.3690	.9651	.3351
.3695	.9653	.3357
.3700	.9656	.3364
.3705	.9659	.3370
.3710	.9661	.3376
.3715	.9664	.3382
.3720	.9667	.3388
.3750	.9669	.3394
.3730	.9672	.3401
.3735	.9676	.3407
.3740	.9677	.3413

.3745	.9680	.3419

Table 10-55
Downcomer
Dimensions.

H/DIA	L/DIA	A_d/A_t
.3750	.9682	.3425
.3755	.9685	.3431
.3760	.9688	.3438
.3765	.9690	.3444
.3770	.9693	.3453
.3775	.9695	.3456
.3780	.9698	.3462
.3785	.9700	.3468
.3790	.9703	.3475
.3795	.9705	.3481
.3800	.0708	.3487
.3805	.9710	.3493
.3810	.9713	.3499
.3815	.9715	.3505
.3820	.9718	.3512
.3825	.9720	.3518
.3830	.9722	.3524
.3835	.9725	.3530
.3840	.9727	.3536
.3845	.9730	.3543
.3850	.9732	.3549
.3855	.9734	.3555
.3860	.9737	.3561
.3865	.9739	.3567

.3865	.9739	.3567
.3870	.9741	.3574
.3875	.9744	.3580
.3880	.9746	.3586
.3885	.9748	.3592
.3890	.9750	.3598
.3895	.9753	.3605
.3900	.9755	.3611
.3905	.9757	.3617
.3910	.9759	.3623
.3915	.9762	.3629
.3920	.9764	.3636
.3925	.9766	.3642
.3930	.9768	.3648
.3935	.9771	.3654
.3940	.9773	.3661
.3945	.9775	.3667
.3950	.9777	.3673
.3955	.9779	.3679
.3960	.9781	.3685
.3965	.9783	.3692
.3970	.9786	.3698
.3975	.9788	.3704
.3980	.9790	.3710
.3985	.9792	.3717
.3990	.9794	.3723
.3995	.9796	.3729
.4000	.9798	.3735
.4005	.9800	.3742
.4010	.9802	.3748
.4015	.9804	.3754
.4020	.9806	.3760

.4020	.9806	.3760
.4025	.9808	.3767
.4030	.9810	.3773
.4035	.9812	.3779
.4040	.9814	.3785
.4045	.9816	.3791
.4050	.9818	.3798
.4055	.9802	.3804
.4060	.9822	.3810
.4065	.9824	.3816
.4070	.9825	.3823
.4075	.9827	.3829
.4080	.9829	.3835
.4085	.9831	.3842
.4090	.9833	.3848
.4095	.9835	.3854
.4100	.9837	.3860
.4105	.9838	.3867
.4110	.9840	.3873
.4115	.9842	.3879
.4120	.9844	.3885
.4125	.9846	.3892
.4130	.9847	.3898
.4135	.9849	.3904
.4140	.9851	.3910
.4145	.9853	.3917
.4150	.9854	.3923
.4155	.9856	.3929
.4160	.9858	.3936
.4165	.9860	.3942
.4170	.9861	.3948
.4175	.9863	.3642

.4175	.9863	.3942
.4180	.9865	.3961
.4185	.9866	.3967
.4190	.9868	.3973
.4195	.9870	.3979
.4200	.9871	.3986
.4205	.9873	.3992
.4210	.9874	.3998
.4215	.9876	.4005
.4220	.9878	.4011
.4225	.9879	.4017
.4230	.9881	.4023
.4235	.9882	.4030
.4240	.9884	.4036
.4245	.9885	.4042
.4250	.9887	.4049
.4255	.9888	.4055
.4260	.9890	.4061
.4265	.9891	.4068
.4270	.9893	.4074
.4275	.9894	.4080
.4280	.9896	.4086
.4285	.9897	.4093
.4290	.9899	.4099
.4295	.9900	.4105
.4300	.9902	.4112
.4305	.9903	.4118
.4310	.9904	.4124
.4315	.9906	.4131
.4320	.9907	.4137
.4325	.9908	.4143
.4330	.9910	.4149

.4330	.9910	.4149
.4335	.9911	.4156
.4340	.9912	.4162
.4345	.9914	.4168
.4350	.9915	.4175
.4355	.9916	.4181
.4360	.9918	.4187
.4365	.9919	.4194
.4370	.9920	.4200
.4375	.9922	.4206
.4380	.9923	.4213
.4385	.9924	.4219
.4390	.9925	.4225
.4395	.9927	.4232
.4400	.9928	.4238
.4405	.9929	.4244
.4410	.9930	.4251
.4415	.9931	.4257
.4420	.9932	.4263
.4425	.9934	.4270
.4430	.9935	.4276
.4435	.9936	.4282
.4440	.9937	.4288
.4445	.9938	.4295
.4450	.9939	.4301
.4455	.9940	.4307
.4460	.9942	.4314
.4465	.9943	.4320
.4470	.9944	.4326
.4475	.9945	.4333
.4480	.9946	.4339
.4485	.9947	.4345

.4485	.9947	.4345
.4490	.9948	.4352
.4495	.9949	.4358
.4500	.9950	.4364
.4505	.9951	.4371
.4510	.9952	.4377
.4515	.9953	.4383
.4520	.9954	.4390
.4525	.9955	.4396
.4530	.9956	.4402
.4535	.9957	.4409
.4540	.9958	.4415
.4545	.9959	.4421
.4550	.9959	.4428
.4555	.9960	.4434
.4560	.9961	.4440
.4565	.9962	.4447
.4570	.9963	.4453
.4575	.9964	.4460
.4580	.9965	.4466
.4585	.9965	.4472
.4590	.9966	.4479
.4595	.9967	.4485
.4600	.9968	.4491
.4605	.9969	.4498
.4610	.9970	.4505
.4615	.9970	.4510
.4620	.9971	.4517
.4625	.9972	.4523
.4630	.9973	.4529
.4635	.9973	.4536
.4640	.9974	.4542

.4640	.9974	.4542
.4645	.9975	.4548
.4650	.9975	.4555
.4655	.9976	.4561
.4660	.9977	.4567
.4665	.9978	.4574
.4670	.9978	.4580
.4675	.9979	.4568
.4680	.9979	.4593
.4685	.9980	.4599
.4690	.9981	.4606
.4695	.9981	.4612
.4700	.9982	.4618
.4705	.9983	.4625
.4710	.9983	.4631
.4715	.9984	.4637
.4720	.9984	.4644
.4725	.9985	.4650
.4730	.9985	.4656
.4735	.9986	.4663
.4740	.9986	.4669
.4745	.9987	.4675
.4750	.9987	.4682
.4755	.9988	.4688
.4760	.9988	.4695
.4765	.9989	.4701
.4770	.9989	.4707
.4775	.9990	.4714
.4780	.9990	.4720
.4785	.9991	.4726
.4790	.9991	.4733
.4795	.9992	.4739

.4795	.9992	.4739
.4800	.9992	.4745
.4805	.9992	.4752
.4810	.9993	.4758
.4815	.9993	.4765
.4820	.9994	.4771
.4825	.9994	.4777
.4830	.9994	.4784
.4835	.9995	.4790
.4840	.9995	.4796
.4845	.9995	.4803
.4850	.9995	.4809
.4855	.9996	.4815
.4860	.9996	.4822
.4865	.9996	.4828
.4870	.9997	.4834
.4875	.9997	.4841
.4880	.9997	.4847
.4885	.9997	.4854
.4890	.9998	.4860
.4895	.9998	.4866
.4900	.9998	.4873
.4905	.9998	.4879
.4910	.9998	.4885
.4915	.9999	.4892
.4920	.9999	.4898
.4925	.9999	.4905
.4930	.9999	.4911
.4935	.9999	.4917
.4940	.9999	.4924
.4945	.9999	.4930
.4950	1.0000	.4936

.4955	1.0000	.4943
.4960	1.0000	.4949
.4965	1.0000	.4955
.4970	1.0000	.4962
.4975	1.0000	.4968
.4980	1.0000	.4975
.4985	1.0000	.4981
.4990	1.0000	.4987
.4995	1.0000	.4994
.5000	1.0000	.5000

The column diameter D is:

(10-567)

The number of tray passes required is determined from Table 10-56.

Table 10-56 Determining the Number of Tray Passes.

Tray spacing (inch)	Increase in number of passes if gpm/in. of weir exceeds
12	3
18	18
24	13

Figure 10-148 and Figure 10-149 show the dimensions of the various downcomers and the distribution area for the number of passes.

Figure 10-148

Dimensions of downcomer types. By permission of Nutter Engineering.

Figure 10-149

Area distribution for different pass types. By permission of Nutter Engineering.
NOTES:

(1) Trays are designed using as few passes as possible. Increasing number of passes decreases efficiency and increases cost.
(2) Multipass trays are designed with equal-bubbling areas. Weir lengths or heights can be adjusted to provide uniform tray hydraulics.
(3) Vapor equalizers are a recommended option to guarantee vapor equalization between compartments.

10.52.1. Diameter of Sieve/Valve trays (F Factor)

The superficial vapor velocity above the tray not occupied by downcomers, is:

(10-568)

where F is the factor based on free column crosssectional area (total column area minus the downcomer area). For foaming systems, the F-factor should be multiplied by 0.75. Figure 10-150 shows the F factor correlation as a function of column pressure for various tray spacings

[284].

Figure 10-150

F Factor as a function of column pressure drop and tray spacing.

(Source: Branan, C., "Rules of Thumb for Chemical Engineers", 4th Ed., Gulf Publishing Professional, 2005).

Branan [284] developed an equation for the F-factor as follows:

(10-569)

Correlation ranges are:

For estimating the downcomer area, Branan has developed an empirical equation to fit Frank's Figure 10-151 [284] as:

(10-570)

Figure 10-151

Estimation of downcomer area for a tray-type distillation column.

(Source: Branan, C., "Rules of Thumb for Chemical Engineers", 4th Ed., Gulf Publishing Professional, 2005).

Clear liquid velocity (ft/s) through the downcomer is then found by multiplying D_L by 0.00223. The correlation is not valid if (very high pressure systems). For foaming systems, D_L should be multiplied by 0.7. Frank recommends segmental downcomers of at least 5% of total column crosssectional area, regardless of the area obtained by this correlation. For final design, complete tray hydraulic calculations are required. For even faster estimates, Branan proposed the following rough F factor guidelines.

Situation	F factor
Fractionating column total cross section vapor velocity	1.0 – 1.5
Sieve tray hole velocity to avoid weeping	>12
Disengaging equipment for separating liquid droplets from vapor	<6

10.52.2. Diameter of Sieve/Valve trays (Lieberman)

Lieberman [284] provides two rules of thumb for troubleshooting fractionators, which could also be used as checks on a design. The pressure drop (ΔP) across a section of trays must not exceed 22% of the space between the tray decks, to avoid incipient flood. This is expressed by:

(10-571)

where

ΔP = Pressure drop in inches of water.
SG = Specific gravity of the liquid on the tray at the appropriate temperature.
T_n = Number of trays
T_s = Tray spacing, in

For sieve trays, a spray height of 15 in. is obtained when the jetting factor is 6 – 7.

(10-572)

where
U = Hole vapor velocity, ft/s
ρ_L = Liquid density
ρ_V = Vapor density

For a 15-inch spray height, a tray spacing of at least 21 inch is recommended.

Example 10-57: Column Sizing Using Nutter's Valve Tray

Size an absorber tower for a methyl ethyl amine (MEA) system with the following conditions:

Vapor flow, lb/h	40,000
Vapor density, lb/ft^3	0.295
Liquid flow, lb/h	330,000
Liquid density, lb/ft^3	61.85
Foam factor	0.75
Residence time, s	4.5
Tray spacing, in.	24
Surface tension, dyne/cm.	57.6

Solution

A computer program PROG105 has been developed to determine the parameters required for sizing a tower using valve trays. The program uses Nutter's Float Design Manual and calculates the bubbling area, downcomer area, tower area and diameter. The tower diameter may be rounded to an appropriate value and the area recalculated. Once the tower diameter is set, the downcomer area ratio is determined and the downcomer height is then computed from Table 10-52Table 10-53Table 10-54 and Table 10-55. Table 10-57 shows the input data and results for tower sizing using valve trays a tower area of 26.57 ft2 and a diameter of 5.82 ft.

Table 10-57 Input Data and Computer Output for Tower Design Using Valve Trays.

Data name: Data105.Dat		
40000.0	330000.0	61.85
0.295	0.75	4.5
24.0	57.6	
Tower Design for Value Trays		
Vapor flow rate, 1b/hr.:	40000.0	
Vapor volumetric rate, ft^3/sec.:	37.665	
Liquid flow rate, 1b/hr.:	330000.0	
Liquid volumetric rate, ft^3/sec.:	1.482	
Liquid density, 1b/ft^3. :	61.850	
Vapor density, 1b/ft^3. :	0.295	
Foam factor:	0.750	
Downcomer residence time, sec.:	4.500	
Tray spacing, in.:	24.	

Surface tension, dyne/cm.:	57 .60
Surface tension/vapor density (X):	195.254
Safety factor:	0.745
Density radical:	0.069
Bubbling area, ft^2:	10.905
Downcomer area, ft^2:	4.446
Tower area, ft^2:	26.571
Tower diameter, ft:	5.816
Downcomer area ratio:	0 .2246

Tray Geometry Sizing

From the tower size (normally obtained by increasing the diameter to the nearest 6-inch increment), the following procedure is used to calculate the downcomer dimensions.

The downcomer area ratio is:

(10-573)

The calculated tower diameter is increased to 6.0 ft. and the tower area becomes:

= 28.27 ft^2

Using Table 10-52Table 10-53Table 10-54 and Table 10-55 and R_{DCA} will allow direct computation of chord height for the side downcomer.

This is rounded up to the nearest ½ in. increment dimension. Using this dimension, the corrected downcomer area is calculated from Table 10-52Table 10-53Table 10-54 and Table 10-55. The computed output of the tower used by the downcomer area ratio (R_{DCA}) is 0.2246.

That is

H/DIA	L/DIA	
0.2760	0.8940	0.2246

The corresponding downcomer height to tower diameter ratio is 0.2760.

The downcomer height =

= 19.87 in. + ½ in.

= 20.4 in.

The new H/DIA = 20.4/72

= 0.2833

This gives a downcomer area to tower area ratio of 0.2329.

H/DIA	L/DIA	
0.2830	0.9009	0.2326

Avg 0.28325	Avg 0.90115	Avg 0.2329
0.2835	0.9014	0.2332

That is:

The downcomer area

= 6.584 ft^2

The weir height, L = (0.90115 × 72)

= 64.88 in (65 in.)

The computed liquid flow rate is:

= 1.483 ft^3/s (665.35 US gpm)

The liquid flow per inch of weir = 665.35/65 = 10.24 US gpm per inch of weir.

A single-pass tray is selected.

10.53. Troubleshooting, Predictive Maintenance and Controls for Distillation Columns

Troubleshooting currently is now much more sophisticated due to the technical tools that are available for investigating and analyzing performance. Radiation scanning referred to as distillation column scanning or “gamma scanning” is now in common use. This is the

process of studying the process material inside a vessel by placing a moving radioactive source inside it, and moving radiation detector along the exterior of the vessel (Figure 10-152a). Such scanning enables the engineer to study tray or packing hydraulics inside the vessel at any set of online conditions as it provides essential data to:

- Track the performance – deteriorating effects of fouling, foaming, flooding and weeping.
- Optimize the performance of vessels.
- Identify maintenance requirements in advance of scheduled turnarounds
- Extend vessel run times.

Figure 10-152a Distillation diagnostics.

(Source: Tru-Tec Division, Koch Engineering Co. Inc.).

Figure 10-152b shows how the radiation source and the detector are aligned on opposite sides of a column or vessel. There are two typical alignments: one-pass trayed column and a packed tower respectively. The source and the detector are synchronized and lowered down the column, while intensity measurements are recorded at specified intervals. The variation in measured intensity depends on the changes in “absorber” thickness and density. The absorber is a composite of the vessel walls, insulation, column internal trays, packings, downcomers, support grids, gas/liquid distributors and process material inside the column. Thus, the difference in intensity that is detected is the result of process changes such as density differences between liquid and vapor within the column.

Figure 10-152b Gamma ray scanning of a single-pass tray and of a packed tower.

The scan produces a density profile that shows the relative density vs. column elevation. The results are then interpreted to show the behavior of the column, its performance and physical condition. Figure 10-152c illustrates how scan results are used to determine tray loading, extent of flooding and vapor space conditions. Tray locations and sources of external interference (e.g. welds, stiffening, rings or supports) are on the left side of the plot.

Flooding: Different extents of flooding occur on tray -1 in the region where scan profile flattens out and rises up.

Clear vapor region: Is marked by the bar line on the right side of the plot. It defines statistically what the relative density should be when only vapor is present.

Froth height intensity: The dotted line to the left of the clear vapor bar is labeled as froth height intensity. This is also referred to as the datum line through the statistical average of the relative vapor space densities and the liquid densities appearing on the tray active areas.

Froth height: Is the vertical distance between the deck and spray height level. This is depicted on tray - 2 on the right side of the vapor bar (Figure 10-152c). The tray is sustaining approximately 9 in. (229 mm) (±1in) of aerated froth.

Clear vapor height: Is the tray spacing minus froth height.

Figure 10-152c Gamma ray scanning profile of a distillation column. The profile is presented as a plot of time (and tray no.) vs. intensity of absorbed emission. The solid line depicts the actual scan data: the clear vapor region is marked by bar line. The dotted vertical lines to the left of the clear vapor bar represent froth height intensity. It is also called datum line.

(Source: Bowman, J.D., Chem. Eng. Prog., Feb. 1991).

Experience with the techniques is important for its effective use combining it with a computer model can assist in both theoretical and practical design applications; best results are obtained by combining scan results and tray packing design [158, 182] it provides data that can aid significantly in determining whether a column is having liquid/vapor flow and or distribution problems. This system provides an accurate density profile of the operating fluids on each tray or through the packing of a packed column, which can be used to identify a number of column malfunctions, as shown in Table 10-58. Figure 10-152d shows an example of using gamma scanning in a depropanizer column to evaluate its performance.

Table 10-58 Column Malfunctions Identifiable by Scanning.

Type of Malfunctions	Particulars
Process	Liquid hold up on tray/packing due to fouling or plugging Foaming Subcooled or superheated feed or reflux
Rate Related	Entrainment – slight, moderate, severe jet flooding Flooding Tray starvation Weeping

Mechanical	Damage to trays and packings Tray displacement Damage to trays due to corrosion Dislocation of liquid or vapor distributors Missing tray manways Level control problems in the base of the column

Figure 10-152d Gamma ray scanning "diagnostic diagnosis" of depropanizer column to evaluate performance. Used by permission, Tru-Tec Division, Koch Engineering Co. Inc.

Other troubleshooting techniques include computer modeling, checking the reliability of instrumentation, measuring quality of product streams with varying reflux rates, measuring column tray temperatures at close intervals, stabilizing the feed rate, bottoms withdrawal and overhead condensing rates. Some surprising problems can be identified, including:

1. Trays may have damage to caps, valves, distributors, sieve holes, or packing for packed towers.
2. The contacting devices of (1) above may actually be missing, i.e., blown off one or more trays, so all that exists is a "rain-deck" tray with no liquid-vapor contact.
3. Crud, polymer, gunk and other processing residues, plus maintenance tools, rags, or overalls may be plugging or corroding the liquid flow paths.
4. Entrainment.
5. Weeping of trays, or flooding of packing or trays.
6. Foaming limitations.
7. Unusual feed conditions, unexpected or uncontrolled.

8. Many other situations, almost too odd to imagine. References on this topic include 159–166, 182, 238.

The topic of control of distillation columns has been discussed by many authorities with a wide variety of experience [117-120, 237], and is too specialized to be covered in this text.

10.54. Distillation Sequencing with Columns having More than Two Products

In the separation of multicomponent mixtures (more than two) using simple columns, there are only two possible sequences (see Figure 10-153). The sequence shown in Figure 10-153 a is known as the direct sequence, in which the lightest component is taken overhead in each column. The indirect sequence as shown in Figure 10-153 b. This removes the heaviest component as bottom product in each column. If we first consider the characteristics of simple columns, a single feed is split into two products. one alternative to two simple columns is shown in Figure 10-154. Here, three products are taken from one column. The designs are feasible and cost effective when compared with simple arrangements utilizing reboilers and condensers operating on utilities for certain ranges of conditions. Where the feed is dominated by the middle product (i.e. >50 % of the feed) and the heaviest product is present in small quantities (i.e. <5 %), then the arrangement shown in Figure 10-154 a can be an attractive option. The heavy product must find its way down the column past the sidestream. If the heavy product has a small flow and the middle product a high flow, a reasonably pure middle product cannot be achieved. In these situations, the sidestream is usually taken as a vapor product to obtain a reasonably pure sidestream.

Figure 10-153

Alternative column sequences for a three-component separation.

Figure 10-154

Distillation columns with three products.

(Source: Smith and Linnhoff, Trans IChemE, ChERD. 66: 195, 1998).

If the feed is dominated by the middle product (i.e. >50%) and the lightest product is present in small amounts (i.e. <5 %), then the arrangement shown in Figure 10-154 b can be an attractive choice. Here, the light product must find its way up the column past the sidestreams. Unless the light product is a small flow and the middle product a high flow, a reasonably pure middle product cannot be achieved. In this instance, the sidestreams is taken as a liquid product to obtain a reasonably pure sidestream [313].

Generally, single-column sidestream arrangements can be attractive when the middle product is in excess and one of the other components is present only at small concentrations. Therefore, the sidestream column only applies to special cases of feed composition. More generally applicable arrangements are made possible by relaxing the restriction that separations must be between adjacent key components.

10.54.1. Thermally Coupled Distillation Sequence

The transfer of heat via direct contact is referred to as thermal coupling, and this is possible when materials flow in such a way as to provide some of the necessary heat transfer by direct contact. In the separation of a multicomponent mixture, thermally coupled distillation columns require less energy and fixed cost compared to conventional multicomponent mixture. To understand this process, first consider the separation of a ternary mixture having components A, B and C. A is the most volatile component and C is the least volatile component, while component B has intermediate volatility. Conventionally, the mixture would be separated by the mixture either via direct sequence or indirect sequence as shown in Figures 10-153a and 10-153b.

For the given ternary mixture, choice between these two arrangements depends on their boiling points, relative volatilities, composition of the ternary mixture, latent heat of vaporization, *etc*. By comparison, thermally coupled distillation columns need less energy and capital investment. Figures 10-155a and 155b show possible arrangements for the separation of the same ternary mixture of A, B and C by thermally coupled distillation. In the thermally coupled side column rectifier, a vapor stream is withdrawn from the stripping section of the main column and sent for rectification in a side column with an overhead condenser. Liquid stream from the bottom of the side column is then returned to the main column.

Figure 10-155

Alternative column sequences for a three-component separation.

In the thermally coupled side column stripper, liquid stream from the main column is withdrawn and returned to the side column for stripping where the reboiler is provided. Vapor stream from the side column is returned to the main column. Selection between these two alternative arrangements depends on concentration profile of intermediate component B. If this reaches a maximum at any point of stripping section of the main or upstream column, and then the side column rectifier is selected. The distillation of B from either C or A is made much easier by withdrawing a side stream from the point at which the concentration of B is maximized.

Figure 10-156 compares conventional and a thermally coupled arrangements in terms of temperature and enthalpy. In the conventional arrangement the pressure of the two columns can be set independently, so allowing variation in temperatures of the two condensers or the two reboilers. Such freedom does not exist in the thermally coupled arrangement. Although thermally coupled arrangement requires a smaller heat load than the conventional arrangement, more of the duties are at extreme levels. These smaller duties favor the heat integration, but the more extreme levels work to its detriment. Therefore, if a thermally coupled arrangement is to be integrated, then the smaller loads and the more extreme levels may work to advantage or disadvantage depending on the situation. [314].

Figure 10-156

Relationship between heat load and level in simple and prefractionator sequences.

(Source: Smith and Linnhoff, Trans IChemE, ChERD. 66: 195, 1998).

It is therefore recommended that simple columns should be used as a first option in the design process. Thermal coupling should be considered when the full heat integration context has been established.

An alternative to thermally coupled columns is a novel type of partitioned tray distillation column (also known as a divided-wall column or Petyluk column) for multicomponent systems. Figure 10-157 shows such a column, where a vertical baffle separates the feed location from that where intermediate boiling product is withdrawn. On the feed side of the partition, separation is achieved between the light (A) and heavy (C) fractions, while the intermediate boiling fraction (B) is allowed to migrate to the top and bottom sides of the partition. On the other side of the partition, the light fraction flows down, so intermediate boiling fraction (B) at the desired purity is withdrawn. Such a column has a larger diameter, but overall capital and operating costs are significantly lower (~30%) [313].

Figure 10-157 Partitioned Distillation Column.

10.54.2. Practical Constraints in Sequencing Options

The following constraints limit the design options that are practically feasible [315]:

1.	Corrosive components must be removed early to minimize the use of expensive, corrosion resistant in materials construction.

2.	Reactive and heat sensitive components must be removed early to prevent product degradation.
3.	Inhibitors are added to compounds that tend to polymerize when distilled. These tend to be nonvolatile, ending up in the column bottoms, and thus prevent finished products from being taken from the bottoms of columns.
4.	Safety considerations often dictate that a particularly hazardous component be removed as early as possible from the sequence in order to minimize the inventory of that material.
5.	When components in the feed to a distillation sequence are difficult to condense, total condensation of these components might require low-temperature condensation employing refrigeration and/or high operating pressures, and hence increased operating costs. Therefore, the light components are normally removed from the top of the first column to minimize the use of refrigeration and high pressures in the overall sequence.

10.54.3. Choice of Sequence for Distillation Columns

Many heuristics/rules of thumb have been proposed for the selection of the sequence for distillation columns, and these are summarized by the following [315]:

10.55. Heat Integration of Distillation Columns

Generally, the scope for integrating conventional distillation columns into an overall process is limited due to practical constraints. The grand composite curve is the curve where enthalpy residuals are displayed as a function of the interval temperatures, and the enthalpy residuals corresponding to the highest and lowest interval temperatures are the minimum heating and cooling utility duties (Figure 10-158). This curve provides a quantitative tool to access the integrating viabilities of the column. If the column cannot be integrated with the rest of the process or if the potential for integrating is limited by the heat flows in the background process, then the distillation process and its complex arrangements should be reviewed.

Figure 10-158

The hot composite curve (HCC) and cold composite curve (CCC) respectively show the heat availability and heat requirement for the overall process.

(Source: Shenoy, Udah, V., Heat Exchanger Network Synthesis: Process Optimization by Energy and Resource Analysis, Gulf Publishing Co., Houston, 1995).

The pinch in this context is the point of minimum temperature difference, representing a bottleneck in heat recovery (i.e. the point of closest approach of composite curves in a “heating and cooling”) problem. It divides the process into two thermodynamically separate regions, above which only hot utility is required, and its magnitude corresponds to the overshoot of the cold composite curve (CCC). Below the pinch, only cold utility is necessary and its magnitude is given by the overshoot of the hot composite curve (HCC). Where the two composite curves overlap, the hot process streams are in enthalpy balance with the cold process streams. The appropriate placement of distillation columns is when heat integration is not across the pinch. In an inappropriately placed column, if it is shifted above the pinch by an increase in pressure, the condensing stream which is a hot stream is shifted from below to above the pinch. The reboiler stream, which is a cold stream stays above the pinch. If the inappropriately placed column is shifted below the pinch by decreasing its pressure, then the reboiling stream, which is a cold stream is shifted from above to below the pinch. The condensing stream stays below the pinch, and therefore appropriate placement is a case of shifting streams, which is a particular case of the plus/minus principle indicated by Smith and Linnhoff [314].

However, when a distillation column is inappropriately placed across

the pinch, its pressure may be altered in order to achieve appropriate placement. However, as the pressure changes, the shape of the “box” changes, since not only do the reboiler and condenser temperatures change but also the difference between the two. This affects the relative volatility, generally decreasing with increasing pressure and subsequently both the height and width of the box will change as the pressure changes.

Changes in pressure also affect the heating and cooling duties for column feed and products. Therefore, the shape of the grand composite curve also changes as the column pressure changes. These effects will not be significant in most processes, since the sensible heat loads involved are small in comparison with the latent heat changes in condensers and reboilers. Figure 10-159 and Figure 10-160 respectively show distillation columns that fit and do not fit against the grand composite curve. See Smith for detailed explanation [315].

Figure 10-159

Distillation columns that fit against the grand composite curve.

(Source: Smith, R. and B. Linnhoff., Trans. IChemE ChERD, 66, 195, 1988).

Figure 10-160

Distillation columns that do not fit against the grand composite curve.

(Source: Smith, R. and B. Linnhoff., Trans. IChemE ChERD, 66, 195, 1988).

If the distillation column will not fit either above or below the pinch, then other design options should be considered. One possibility is splitting the column feed and feeding to two separate parallel columns. The relative pressures in the columns are chosen such that the two columns can each be appropriately placed. The capital cost of such a scheme will be higher than that of a single column [314]. Use of an intermediate reboiler or condenser can also be considered. Reviews of this approach are presented elsewhere [316, 317, 318].

10.56. Capital Cost Considerations for Distillation Columns

Separators (e.g. distillation columns, evaporators and dryers) are energy intensive. Their efficiency in terms of the overall process can be improved if they are properly heat integrated. In distillation, the major design parameters must be ascertained to allow the design to proceed. The first decision is the operating pressure. As this is raised:

1. Condenser temperature increases.
2. Vapor density increases, resulting in a smaller column diameter.
3. Reboiler temperature increases with a limit often set by thermal decomposition of the material being vaporized. This results in excessive fouling.
4. Latent heat of vaporization decreases, *i.e.* reboiler and condenser duties become lower.
5. Separation becomes more difficult (relative volatility, α decreases), *i.e.* more plates or reflux are required.

Correspondingly, as the operating pressure is lowered these effects are reversed, and the lower limit is often set by the desire to avoid:

- Refrigeration in the condenser

- Vacuum operation.

The use of refrigeration and vacuum operation incurs further capital and operating costs and thus increase the complexity of the design. It is essential, therefore to set distillation pressure to as low a pressure above ambient as allows cooling water or air cooling to be used in the condenser. The pressure should be fixed such that the bubble point of the overhead product is 10°C above the summer cooling water temperature, or to atmospheric pressure if using vacuum operation. When distilling high molecular weight material, process constraints dictate that vacuum operation be used in order to reduce the boiling point of the material to below where product decomposition occurs.

Reflux ratio is also needs to be chosen for distillation. Figure 10-161 illustrates a capital-energy tradeoff in a stand-alone distillation column. As the reflux ratio increases from its minimum, the capital cost decreases initially as the number of plates reduces from infinity, but the utility costs then increase as more reboiling and condensation are required. The optimal ratio of actual to minimum reflux is often less than 1.1, but designers are reluctant to design columns closer to minimum reflux than 1.1 except in special circumstances. This is because a small error in design or small alteration in operating conditions might result in an infeasible design. If the column is properly heat integrated with the rest of the process, the reflux ratio can often be increased without changing the overall energy consumption, as illustrated in Figure 10-162 a. Increasing the heat flow through the column decreases the number of plates required, but increases the vapor rate. The corresponding decrease in heat flow through the process (shown in Figure 10–162 b) will have the effect of decreasing temperature driving forces and increasing the capital cost of the heat exchanger network. Therefore, the tradeoff for an appropriately integrated distillation column becomes one between the

capital costs of the column and the capital cost of the heat exchanger network-as shown in Figure 10-162 c.

Figure 10-161

The capital energy tradeoff for stand-alone distillation columns.

Figure 10-162

The capital/capital tradeoff for an appropriately integrated distillation column.

(Source: Smith and Linnhoff, Trans IChemE, ChERD. 66: 195, 1998).

The nature of the tradeoffs change and the optimal reflux ratio for the heat-integrated column can be very different from that for a stand-alone column and thus the optimal reflux ratio for an appropriately integrated distillation column will be problem-specific.

For more examples, please visit: http://www.elsevierdirect.com/companions/9780750683661

Nomenclature for Part 1: Distillation Process Performance

A, B thru K = Constants developed in original article

a, b, c = Correlation constants (distillation recoveries [141])

= constants in

generalized Antoine equation (Table 10-5)
a = Activity of component
a_i = Activity of component, i
av or avg = Average
= attraction function in SRK equation 10-75,
= constant value of at the critical point,
B, C, D = Virial coefficients, Equation 10-11
B = Bottoms product or waste, lb mols/hr, also = W
B_b = Mols of component, b, used as reference for volatility, after a given time of distillation
B_{bo} = Mols of component, b, used as reference for volatility, at start of distillation
B_i = Mols of component, i, after a given time of distillation
B_{io} = Mols of component, i, at start of distillation
B_{Ti} = Total mols of liquid in bottoms of still at time, T_1
B_{To} = Total mols liquid (not including any steam) in bottom of still at start time T_o (batch charge)
b = y intercept of operating line; or constant at fixed pressure for Winn's relative volatility
b = intercept of a straight line with the y-axis
b_i = Mols of component, i, in bottoms
= volume parameter in SRK equation 10-76,
.
C = No. components present, phase rule; or no. components, or constant

C_{mi} = Factor in Colburn Minimum Reflux method, pinch conditions, stripping

C_{ni} = Factor in Colburn Minimum Reflux method, pinch conditions, rectifying

C = Specific heat, Btu/lb (°F)

C_{mi} = Mols of distillate or overhead product, lb mols/hr; or batch distillation, mols

C_p = heat capacity, Btu/(lb mole °F)

c = constant defined by Equation 10-253

D = moles distillate per unit time

d_i = Mols component, i, in distillate

E = Vaporization efficiency of steam distillation

E_G = Overall column efficiency

E_o = Overall tray efficiency

E_{MV}* = Eoc = Murphree point efficiency, fraction

E_{MV}O = Murphree plate/tray efficiency, = EM

F = Degrees of Freedom, phase rule; or, charge to batch still, mols

F = Feed rate to tower, lb mols/hr; or, mols of feed, (batch distillation) entering flash zone/time all components except noncondensable gases

F_{FR} = Factor for contribution of other feed flow to minimum reflux

F_L = Mols of liquid feed

F_v = Mols of vapor feed

$F_t = F + V_s$ = mols feed plus mols of noncondensable gases

$F_R = F_{SR,k}$ = Factor for contribution of side stream, k, flow to minimum reflux

F_{SR} = Factor for contribution of sidestream flow, to minimum reflux

F_{FRJ} = Factor for contribution of feed, j, flow to minimum reflux

f = Fugacity at a specific condition

f = acentric factor function, [illegible] in Table 10-5, [illegible] in Table 10-6, dimensionless

[illegible] = acentric factor dimensionless in Equation 10-77.

[illegible] = acentric factor function for SRK equation, dimensionless.

[illegible] = acentric factor function for PR equation, [illegible] dimensionless

fυ = Fugacity at reference standard condition

f_i = Feed composition,. i,; or, = total mols of component, i, in distillate and bottoms

G = Boilup rate, mols/hr

H = Total enthalpy, above reference datum, of vapor mixture at tray or specified conditions, Btu/lb mol, or Btu/lb

H′ = H_{ij} = Henry's Law constant, lb mols/(cu ft) (atm)

H_n = Total molal enthalpy of vapor at conditions of tray, n, entering tray; $H_n = \Sigma H_{ni} (y_{ni})$

H_s = Total enthalpy of steam, Btu/lb mol, or Btu/lb

HK = Heavy key component in volatile mixture

h = Enthalpy of liquid mixture or pure compound at tray conditions of temperature and pressure, or specified point or condition, Btu/lb mol, or Btu/lb

h = ratio of parameter [illegible] to molar volume, Equation 10-81, dimensionless

h_n = Total molal enthalpy of liquid at conditions of tray, n; h'_n =

$h_{ni}(x_{ni})$

h_D = Molal enthalpy of product or total liquid enthalpy above reference datum for sum of all contributing percentages of individual components

h_{n+1} = Molal enthalpy of liquid leaving plate n + 1

h_a = Total molal enthalpy of liquid at conditions of tray, n; hn = Σh_{ni} (x_{ni})

K = Equilibrium constant for a particular system (= y/x)

K′ = Equilibrium constant for least volatile component, K′ = y/x

K_i = Equilibrium distribution coefficient for component, i, in system

k = Experimentally determined Henry's Law constant, also can be K

k = Value of x at intersection of operating line and equilibrium curve on x–y diagram (batch operation)

k = composition of the liquid where the operating line intersects the equilibrium line

kpa = Metric pressure

L = Liquid flow rate return to tower as reflux, lb mols/hr, mols component in liquid phase; or, L_1, L_2 = Latent heat of vaporization; or, volumetric flow rate for incoming contaminated water (stripping VOC with air); or mols liquid produced from F per unit time, leaving flash zone, or moles liquid per unit time, rectifying section

= moles liquid per unit time, stripping section.

L_F = moles liquid from feed plate per unit time.

L_o = moles reflux liquid returned to plate 1 per unit time.

L_r = Liquid flow rate down rectifying section of distillation tower, lb mols/hr

L_s = Liquid flowrate down stripping section of distillation tower,

lb mols/hr
LK = Light key component in volatile mixture
L/V = Internal reflux ratio
L/D = Actual external reflux ratio
$(L/D)_{min}$ = Minimum external reflux ratio
M = Molecular weight of compound
M_s = Total mols steam required
m = Number of sidestreams above feed, n
N = Number of theoretical trays in distillation tower (not including reboiler) at operating finite reflux. For partial condenser system N includes condenser; or number theoretical trays or transfer units for a packed tower (VOC calculations) N_B = Number of trays from tray, m, to bottom tray, but not including still or reboiler
N_{min} = Minimum number of theoretical trays in distillation tower (not including reboiler) at total or infinite reflux. For partial condenser system, N_{min} includes condenser; also, minimum value of N
N_n = Number of theoretical trays above feed, or reference plate, n, but not including n
N_m = Number of theoretical trays before feed tray
N_{im} = Mols of immiscible liquid
N_o = Mols of nonvolatile material present; or, number of theoretical trays/stages in column only, not reboiler or condenser
N_s = Mols of steam
n = Number of theoretical trays in rectifying section or number of components, or minimum number of equilibrium trays
n = number of plates between any two points in the column under consideration.
n_f = Number of feeds
n_s = Number of sidestreams
P = Pressure, atmospheres; or, vapor pressure of component,

atm.; or,

P = number of phases; or, P = for batch operations, percentage drawoff

P_i = Vapor pressure of each component

P_s = Vapor pressure of steam, absolute

P_b = Vapor pressure of reference more volatile component, b

p = Pi = Partial pressure of one compound in liquid, absolute units, or, ratio rps/rpr; also, P_i = partial pressure of solute (Henry's Law)

p = Total pressure of system = π

p′ = Number of trays below feed where introduction of light components should begin, Akers-Wade calculation method

P_i^* = Vapor pressure component, i, in pure state at temperature

P_{ii}^* = Similar to above by analogy

p″ = Number of trays above feed where introduction of heavy components should begin. Akers-Wade calculation

p_{im} = Pure component vapor pressure of immiscible liquid, mm Hg

p_s = Partial pressure of steam, mm Hg

P_t = total pressure.

Q_B = Net heat in through reboiler duty, Btu/hr; or heat added in still or bottoms

Q_c = Net heat out of overhead condenser, Btu/hr, = w cp $(t_i - t_o)$

q = qF = Thermal condition of feed, approximately amount of heat to vaporize one mol of feed at feed tray conditions divided by latent heat qf vaporization of feed

q_B = reboiler duty, Btu/hr.

q_D = condenser duty, Btu/hr.

q_v = Thermal condition of sidestream (s)

R = Reflux ratio = External reflux ratio for a given separation, = L/D, L = liquid rectifying column

R = reflux ratio, reflux per unit time divided by moles of distillate per unit time

R = Actual reflux ratio, O/D

R_m = Minimum reflux ratio, O/D

R′ = Pseudo minimum reflux ratio

R_{min} = Minimum external reflux ratio for a given separation

R_F = Feed component of minimum reflux

$R_{F,n}$ = Feed component of minimum reflux for feed, n

R_{OF} = Other feed components of minimum reflux

R_s = Sidestream component of minimum reflux

r_{ps} = Ratio of light to heavy keys, stripping pinch

r_{pr} = Ratio of light to heavy keys, rectifying pinch

r_f = Ratio mol fraction light key to heavy key in feed

S = Steam flowrate, lb1lhr or lb mols/hr; or theoretical stages at actual reflux (Figure 10-24a and Figure 10-24b) including reboiler and partial condenser, if any; or batch, mols in mixture in still kettle at time θ

S_n = Theoretical stages at minimum reflux

S_M = Minimum theoretical stages at total reflux from bottoms composition through overhead product composition, including reboiler and any partial condenser (if used); or minimum stripping fat for at minimum flowrate of air

S_k = Flowrate of sidestream, k, mols/hr

S_o = Theoretical stages at a finite operating reflux; or batch, mols originally charged to kettle

S_r = Theoretical stages in total rectifying section, including partial condenser, if used

S_s = Theoretical stages in total stripping section, including reboiler

S_t = Theoretical trays/stages at actual reflux, L/D, including reboiler and total condenser

S_{opt} = Optimum stripping factor
$(SR)_i$ = Separation factor
s = Pounds (or mols) steam per pound (or mol) of bottoms; or flowrate of sidestream, mols/hr
T = Temperature, °Abs R
t_B = Bottoms temperature, °F
t_i = Temperature in, °F
t_o = Temperature out, oF; or overhead temperature, °F
V = V_t = Total vapor leaving flash zone/unit time at specific temperature and pressure; or total overhead vapor from tower, mols/hr; or mols of component in vapor phase; or volumetric flowrate for incoming fresh air
V = Quantity of vapor, mols
V_C = moles of vapor condensed necessary to heat reflux liquid to t.
V_F = moles vapor per unit time rising from feed plate.
V_r = Vapor flow rate up rectifying section of tower, lb mols/hr
V_s = Vapor flowrate up stripping section of tower, lb mols/hr; or mols noncondensable gases entering with feed, F, and leaving with vapor, V/time
V_{min} = Minimum fresh air flow based on slope of operating line, L/V, on x–y diagram
v = Vapor flow rate, mols/hr; or molar volume
W = Bottoms product, or still bottoms, or kettle bottoms, mols; also see B; or mols/hr bottoms product; or mols of residue or bottoms/unit time (Ponchon heat balance)
W = Weight of material in vapor (steam distillation)
W_1 = Mols final content in still
W_{il} = Contents of still pot or kettle at any point, 1, after start for components, i, mols
W_{io} = Initial contents of kettle or still pot, mols, for component, i

W_o = Mols liquid mixture originally charged to still pot
w = Pounds coolant per hour
x_B = mole fraction in the bottoms.
x_D = mole fraction in the distillate.
x_i = x = Mol fraction of component in liquid phase; or mol fraction solute in solution (Henry's Law)
$x_f = x_{lF}$ = Mol fraction of any component in feed, vapor + liquid, F_t; $x_f = Fx_f/F_t$
x′ = Mol fraction of least volatile component
$x'_1 = x_l - k$
$x'_p = x - k$
x = Mol fraction more volatile component in liquid
x_1 = Mol fraction of component, i, in liquid mixtures as may be feed distillate or bottoms, B_T, at any time, T_l; or mol fraction more volatile in vapor entering column at any time (or in distillate)
x_{it} = Mol fraction liquid at intersection of operating lines at minimum reflux, Scheibel-Montross equation
x_{hf} = Mol fraction heavy key in feed
x_n = Pinch composition and light component mol fraction
x_N = Mol fraction VOC component in the stripped water exiting, usually targeted at meeting environmental regulations
x_{iD} = Mol fraction light key component in overhead product; or, any light component Colburn)
x_{iB} = Mol fraction light key component in keys in original charge
x_{io} = Mol fraction light key in overhead expressed as fraction of total keys in overhead
x_{1B} = Mol fraction most volatile component in bottoms
x_{hD} = Overhead composition of heavy key component, mol fraction
x_{hn} = Pinch composition of heavy key component, mol fraction

x_l = Mol fraction of component in liquid phase; or mol fraction more volatile component in vapor entering column at any time

x_s = Mol fraction of a more volatile in kettle at time θ

x_{si} = Value of xs when distillate receiver is first filled

x_{so} = Mol fraction more volatile in kettle at time θ

x_w = Mol fraction more volatile component in bottoms residue (final); or, composition of liquid in still, mol fraction

x_{wo} = Initial mol fraction of more volatile component in liquid mixture

x_F = Mol fraction more volatile component in feed

x_D = Mol fraction more volatile component in final distillate = mol fraction in distillate leaving condenser at time θ

x_p = Mol fraction of more volatile component in liquid leaving column at any time

x_L = Mol fraction of feed as liquid, Scheibel-Montross

x_{lo} = Mol fraction light key in overhead expressed as fraction of total keys in overhead, Scheibel-Montross equation

x_m = Tray liquid mol fraction for start of calculations (most volatile component)

x_o = Mol fraction of component, i, in bottoms B_{To} at start time, T_o; or VOC mol fraction

x_v = Mol fraction of feed as vapor, Scheibel-Montross equation

$y = y_i$ = Mol fraction of component in vapor phase, as may be feed, distillate, or bottoms; or Henry's Law, y_i = mol fraction solute in vapor

y_i = Mol fraction VOC component in the exiting VOC contaminated air

y' = Mol fraction of least volatile component

y^* = Equilibrium value corresponding to x_i

y_n = Average light key mol fraction vapor leaving plate, n

y_{n-1} = Average light key mol fraction vapor entering plate, n + 1
y_{n-1} = Mol fraction VOC component in the incoming fresh air (equals zero for fresh air)
y_j = Mol fraction solvent component in vapor
y_s = Mol fraction steam in vapor
Y_{iB} = Percent recovery of, i, in the bottoms
Y_{iD} = Percent recovery of, i, in the distillate
Z = Compressibility factor
z_F = mole fraction in the feed.
$z_{i,F}$ = Mol fraction component, i, in feed
$z_{i,Fj}$ = Mol fraction component, i, in feed, j
$z_{i,S}$ = Mol fraction component, i, in sidestream
$z_{i,Sk}$ = Mol fraction component, i, in sidestream, k

Greek Symbols

α, α_1 = Relative volatility of light key to heavy key component, or any component related to the heavy key component, except Equation 8-65, α_i is based on heavy key)
α_{avg} = Average relative volatility between top and bottom sections of distillation tower/column
α_j = Relative volatility of more volatile to each of other components (steam distillation)
α_i = Relative volatility of component, i
α_H = Relative volatility of components heavier than heavy key at feed tray temperature
α_i = Relative volatility of more volatile to each of other components
α_L = Relative volatility of components lighter than light key at feed tray temperature
β = Constant of fixed pressure in Wino's relative volatility,

Equation 8-43

θ = Time from start of distillation to fill receiver, or value of relative volatility (Underwood Parameter) to satisfy Underwood Algebraic Method

θ_1 = Time for filling distillate receiver, hrs

θ_2 = Time for refluxed distillation (batch), hrs

μ = Viscosity, centipoise

υ = Activity, coefficient

π = total system pressure, absolute; atm, mm Hg, psia

π = 3.14159

Σ = Sum

Ψ = First derivative function

Ψ' = Second derivative function

ω_j = Function in Underwood's Algebraic method for minimum reflux ratio

Ω = Fugacity coefficient

ϕ = No. phases from phase rule

λ = Latent heat of vaporization, Btu/lb mole.

Subscripts

a, b, c, *etc.* = Specific components in a system or mixture

Avg, Av = Average

B = Any consistent component in bottoms product

B = b = Bottoms

b = Exponent in Winn's relative volatility equation

D = Any consistent component in condensed overhead product or distillate

eff = Effective

F = Feed

F_j = Feed, j

F_n = Intermediate feed, Scheibel-Montross method

F_L = FH = All mol fractions lighter than light key in feed,

Scheibel-Montross method
FHK = Heavy key in feed
FLK = Light key in feed
HK = h = h_k = Heavy key component
H = Components heavier than heavy key
h = Heavy, or heavy or high boiling component in mixture; also heavy key component
i = Any component identified by subscripts 1,2, 3, etc, or by a, b, c, etc.; or initial condition, i
im = Immiscible liquid
j = Specific components in a system or mixture
1 = 1k = Light key component; or light or low boiling component in mixture
1h = Refers to light component referenced to heavy component
LK = Light key component
L = Liquid, Scheibel-Montross method only; or components lighter than light key
M = min = Minimum
m = No. trays in stripping section; or tray number
n = No. trays in rectifying section; or tray number
o = Initial conditions; or i; or operating condition
p_r = Pinch condition in rectifying section
p_s = Pinch condition in stripping section
P = For packed towers
w = Relates to bottoms or pot liquor, or kettle bottoms
r = Rectifying section; or component to which all the relative volatilities are referred
s = Steam, or stripping section of column
t = Top, or total
T = For tray towers
v = Vapor
1 = Initial, steam distillation
2 = Remaining, steam distillation

1, 2, 3, *etc.* = Tray numbers; or specific components in a system or mixture

Chapter 11. Petroleum, Complex-Mixture Fractionation, Gas Processing, Dehydration, Hydrocarbon Absorption and Stripping

Part 2: Fractionation

The demand for crude oil is often market-driven, as recently observed in the world economic recession where the price of U.S crude oil fell from U.S. $140 to $36/barrel by the end of 2008. Crude oil distillation is a large energy consumer and a large waste generator, so refiners need to balance the economics of processing crude oil to viable fractions. The thermodynamics of multicomponent distillation, as discussed earlier in the book, apply to petroleum, synthetic crude oil and other complex mixtures. Various advanced integrated and distributed distillation methods can be employed to minimize waste and thus optimize performance oil [295].

The processing of crude oil requires special consideration for the following reasons:

The feed stocks are generally complex mixtures of many different types of hydrocarbons along with inorganic and organic compounds. The number of carbon atoms in the component molecules ranges from 1 to more than 50, so the compounds have boiling points from –260°F (–162°C) to more than 1000°F (538°C). Further, the number of different compounds of similar volatility increases very rapidly with corresponding increase in boiling point, *e.g.* 16 of the 18 octane isomers boil within a range of only 22°F (12°C).

The fractionation products of these complex mixtures are often complex mixtures themselves, and their character and yield varies widely, and often depends upon the source of the feedstock. Crude

oils from the same locality sometimes exhibit marked variations.

The scale of petroleum fractionation/distillation is generally large, involving atmospheric distillation of crude oil, vacuum distillation of bottoms residue, main fractionation of gaseous effluent from catalytic cracking, and main distillation of effluent from thermal coking of various petroleum fractions. They use heavy equipment which consumes large quantities of energy. Figure 11-1 shows a flow diagram of a typical refinery distillation unit.

Figure 11-1

A flow diagram of a refinery distillation unit.

11.1. Characterization of Petroleum and Petroleum Fractions

The primary raw material for all petroleum processes is crude oil, a naturally occurring liquid produced from wells drilled into the earth. The well depth and techniques used to produce the oil vary widely around the world, as does the nature and chemical composition of the crude oil itself.

The elementary composition of crude oil will, however, fall within the following ranges:

Element	% by weight
Carbon	84–87
Hydrogen	11–14
Sulfur	0.06–8
Nitrogen	0.02–1.7
Oxygen	0.68–1.82

Oxygen	0.68–1.82
Metals	0–0.14

Crude oil is classified according to the family of hydrocarbons it contains, as paraffin base, naphthene base, aromatic base or asphalt base. The refining process is dependent on the crude oil base. The relative values of crude oil depend on the processing costs and the market values of the associated products.

11.1.1. Crude Oil Properties

Crude oil is a mixture of thousands of hydrocarbons; the American Petroleum Institute (API) Project 6 has isolated over 16,000 distinct hydrocarbon compounds from a single sample. The tests outlined in the following sections are employed to characterize crude oils as potential refinery feeds.

11.1.2. Gravity, API

The density of petroleum oil is expressed in the United States as American Petroleum Institute (API) gravity, where API gravity is related to specific gravity as follows:

(11-1)

Both specific gravity and API gravity are expressed relative to the density of water at 60°F (15.9°C). The expression for API is nonlinear, and so specific gravities must be used when calculating the gravity for blends. The API gravity is inversely related to the specific gravity, so a higher value shows a less dense material (e.g. the API of water is 10.0). API gravity is preferred to specific gravity as a measure of density because the specific gravities of various crude oils lie very close to each other and the API gravity alone can differentiate

between them. API gravity is determined with a hydrometer and corrected to 60°F (15.9°C).

In the case of a mixture or blend of various fractions, the API gravity is determined as the sum of the individual API gravities multiplied by the corresponding weight fraction in the mixture.

(11-2)

where

API_1, API_2, API_3, = the respective API for each components

W_1, W_2, W_3, = the respective weight fractions

The specific gravity of crudes can be categorized into four groups [296]:

Light crude oil, Sp Gr: *	<0.825
Medium crude oil, Sp Gr:	<0.875
Heavy crude oil, Sp Gr:	<1.000
Extra heavy crude oil, Sp Gr:	>1.000
* represents density at 15°C with respect to that of water at 4°C.	

Table 11-1 shows typical specific gravities of some crudes.

Table 11-1 Specific gravities of some crude oils

Country	Crude	Sp. gravity:

Indonesia	Minas	0.845
Iraq	Kirkuk	0.845
Iran	Cyrus	0.940
Kuwait	Kuwait	0.870
Nigeria	Bonny Light	0.837
Saudi Arabia	Arabian Light	0.858
Venezuela	Boscan	1.000

11.1.3. Boiling Point Range

The boiling point range for a crude oil is determined with a *true boiling point* (TBP) distillation test. This test is performed in an apparatus with sufficient fractionation to recover samples of the various products that can be distilled from crude oil. Results are reported at one atmosphere. It is impossible to determine the complete boiling point range for most crude oils, as they thermally degrade above temperatures around 650°F (344°C) and a pressure of one atmosphere. Therefore the pressure is gradually reduced during a TBP distillation test to approximately 40 mm Hg for the heaviest fraction. These conditions are equivalent to a temperature at atmospheric pressure of 900–950°F (483–510°C). Nelson and others [297] state that the maximum boiling point for crude oil mixtures is greater than 1400°F (760°C).

11.1.4. Characterization Factor

Watson Factor (K_W)

Watson, Nelson and Murphy [298] proposed the following characterization factor:

(11-3)

where

K = characterization factor

= the mean average boiling point of the sample, °R

Sp. Gr. = specific gravity at 60°F relative to water.

This characterization factor, which is a rough indicator of hydrocarbon class (or composition) is related to molecular weight, viscosity, critical temperature, *etc.* K ≤ 10 indicates an aromatic material whereas K = 15 indicates a paraffinic material. For most crude oils, . Figure 11-2 is a nomograph for estimating the K value from the mean average boiling point, the API gravity and an ASTM D-86 distillation of the petroleum fraction. The characterization factor, K_W is defined in the example given on the figure.

Figure 11-2

Specific Gravity of Petroleum Fractions.

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The Universal Oil Product Characterization factor, K_{UOP}

K_{UOP} is widely used since it expresses quantitatively the variations in physical properties that vary with the character of the stock. It is defined as the cube root of the molal average boiling point, T°R

divided by the specific gravity at 15.6°C.

(11-4)

Where T is the average boiling point in °R and d is the specific gravity at 15.6/15.6°C.

The volume average boiling point (VABP), obtained either from True Boiling Point (TBP) distillation or American Society for Testing and Materials (ASTM) distillation, replaces T in the expression for K_{UOP}.

In the case of a crude oil subjected to TBP distillation (given as volume),

or in the case of a petroleum cut subjected to ASTM distillation:

where T_{20}, T_{50}, T_{80}, T_{90} are the temperatures at which 20, 50, 80 and 90 per cent by volume of a sample distils out.

It is found that highly paraffinic oils have K_{UOP} values from 12.5–13.0 and cyclic (naphthenic/aromatic) oils have K_{UOP} values from 10.5–12.5. Thus K_{UOP} values can differentiate between highly paraffinic and highly aromatic stocks but have problems in classifying naphthenic stocks from aromatic ones, since their K_{UOP} values overlap.

Note: In the case of a mix, the characterization factor is determined as:

(11-5)

(11-5)

where

K_1, K_2, K_3 = the respective characterization factors for different components

W_1, W_2, W_3 = the respective weight fractions

11.1.5. Carbon Residue, wt%

This property is determined by reducing a sample to elemental carbon (coke) in the absence of air. It is a measure of the asphalt content of the crude oil and the quantity of lubricating oil fraction that may be recovered as product. It can also be used to establish the yields from coking processes and the coke deposited on the catalyst in catalytic cracking. The higher the carbon residue for a crude oil, the lower its value as a refinery feed stock.

11.1.6. Nitrogen Content, wt%

Nitrogen compounds in crude oils reduce the activity of cracking catalysts. High nitrogen content crude oil may require special processing to remove the nitrogen. Most of the nitrogen in crude oil is contained in the highest boiling fractions.

11.1.7. Sulfur Content, wt%

Sulfur content and API gravity have the greatest influence on the value of crude oil. It varies from trace levels to >5 wt%. High sulfur oils are referred to as "*sour*" crude oils and require more extensive processing than do low sulfur or "*sweet*" oils. Generally, the lower the API gravity, the higher the sulfur content, although there are exceptions to this rule. Further processing for sour crude oils involves hydrogen treating processes to convert the sulfur to H_2S, which can the be removed.

11.1.8. Salt Content, pounds/1000 barrels

All crude oil contains some inorganic salts, generally in the form of chlorides. If the salt content exceeds 10 pounds/1000 barrels, it is necessary to remove them prior to distillation to reduce corrosion in the distillation column and its overhead condensers.

11.1.9. Metals, parts/million (ppm) by weight

Many metals occur in crude oil, the most common being nickel and vanadium. Metals tend to concentrate in the heavier fractions where they act as poisons in the catalytic and hydrogen cracking processes. Other metals include copper, zinc, iron and arsenic.

11.1.10. Pour Point (°F or °C)

The pour point (ASTM D97) is used to determine the waxiness of petroleum oil. Waxes are paraffins with long side chains, or long side chains attached to other hydrocarbon structures. A waxy crude oil has a high pour point; *i.e.* it solidifies at a temperature above room temperature. Conversely, a low pour point indicates an absence of paraffinic and/or a presence of aromatic compounds, as these tend to lower the pour point. A sample of oil is first heated to 115°F (46.4°C) to ensure complete liquefication. It is then cooled to 90°F (32.6°C). The sample is then placed in a cooling bath maintained between 15–30°F (8.3–16.7°C) below the pour point of the oil. The oil is removed from the bath at 5°F (2.8°C) intervals and it tilted to determine whether the sample will flow. If no movement is detected within 5 seconds, the solid point has been reached. It is defined as the temperature 5°F (2.8°C) above the solid point.

The same procedure is used to determine the cloud point of petroleum oil. This is the temperature at which a distinct cloudiness or formation of crystals appears in the bottom of the test jar. It is typically about 5°F (2.8°C) higher than the pour point. Table 11-2 gives pour points of some typical crude oils.

Table 11-2 Pour points of some crude oils

Country	Crude	Pour point, °C
Algeria	Hassid Messaoud	−60
Kuwait	Kuwait	−42
Libya	Dahra	−1
Venezuela	Boscan	+15

11.2. Crude Oil Assay Data

A precise and complete analysis of a crude oil is referred to as crude assay. This is more detailed than a TBP curve. Crude oil assay data are available for most of the crude oils of the world; some data are company proprietary, but much is available in open literature. The crude assay helps to assess the potential sales value of a new oil and thus plan for its effective utilization.

The following sections cover the information obtained from an assay.

11.2.1. Whole crude oil average properties

Gravity, weight % sulfur, weight % nitrogen, pour point, viscosities, metal contents, weight percent carbon residue, Reid vapor pressure, salt content, flash point, % asphaltene, bottom sediment and water (BS & W), light hydrocarbon yields, (C_1–C_5), and a boiling point curve are the both properties of the oil given by an assay.

11.2.2. Fractional properties

These are the boiling point or cut-point range and yield on crude, research octanes for gasoline fractions, smoke and freeze points for kerosenes, naphthene and aromatic contents for reforming stock, and cetane number for diesel fuels.

Crude oil assay data change over time for any oil. This is caused by two factors; firstly, the crude oil itself changes in composition, and

secondly, a crude oil sample represents a mixture of the crude oil produced in various wells and fields so, as new one are brought on line, the composition of the mixture changes. It is good practice to request a confirming sample for a crude oil mixture being used in any process design/simulation exercise.

11.3. Crude Cutting Analysis

The first seven crude cuts are pure components. They are the light ends and consist of H_2S, H_2, C_1, C_2, C_3, iC_4, nC_4. The iC_5 and nC_5 are together and therefore are treated as the eight cut with a TBP end point of 100°F. The sum of light ends through nC_5 should be equal to the lights vol % distilled at 100°F. However, if the two numbers do not agree, then the light ends yields are assumed to be correct and the TBP yields between 100–200°F are adjusted to provide a smooth transition between light ends and the remaining TBP curve.

The ninth cut has a TBP cut point of 100–200°F. Each successive cut has a TBP end point of 20°F higher than the preceding component. The 63rd component has end point of 1200°F. The TBP end point of a crude is difficult to estimate with any degree of accuracy in the laboratory and is therefore assumed to be 1292°F. The 64th cut is assumed to have a TBP cut range of 1200–1292°F. Table 11-3 and Table 11-4 show crude assays of Middle East Arab extra light and heavy crudes.

Table 11-3 Middle East: Arabian Extr

Crude Data Sample Date: SG:	**01/06/930.8386**	**API Gravity**	**37.2**	**Analysis Date: Sulfur:**	**1.38 wt%**	
Light Hydrocarbon Analysis						

	wt %	lv%				
Hydrogen sulfide	0.00	0.00				
Methane	0.00	0.00				
Ethane	0.00	0.00				
Propane	0.07	0.12				Total t
Isobutane	0.11	0.16				
N-butane	0.77	0.11	Total to C4s	0.88	1.27	Total t
Isopentane	0.70	0.94				
N-pentane	1.40	0.86	Total to C5s	2.10	2.80	Total t
Wide Cut Data TBP Cut Points	**C**	**C5–95**	**95–175**	**C5–149**	**149–232**	**332**
Yield on Crude	wt%	7.91	14.36	15-75	16.35	21.50
	lv%	9.86	16.03	19.88	17.44	21.47
SG		0.6729	0.7512	0.7068	0.7859	0.8396
API Gravity		78.79	56.87	68.71	48.54	37.04
Watson K factor		12.54	12.03	12.21	11.96	11.94
Sulfur	wt%	0.009	0.028	0.019	0.069	0.921
Mercaptan Sulfur	wt%	0.018	0.29	0.025	0.025	0.026
Iso Paraffins	wt%	37.79	31.18	34.19	30.57	-
Total Paraffins	wt%	86.77	64.88	76.45	58.93	-
Naphthenes	wt%	11.49	19.60	16.15	18.42	-
Aromatics	wt%	1.74	15.52	7.40	22.65	29.27
Viscosity at 20°C	cSt	-	0.87	-	1.39	5.07
30°C	cSt	-	0.77	-	1.20	3.94
50°C	cSt	-	0.62	-	0.93	2.59

80°C	cSt	-	0.47	-	0.68	1.61
100°C	cSt	-	0.40	-	0.56	1.25
Cloud Point	C	-	−80.5	-	−56.7	−12.0
Freeze Point	C	-	−77.3	−96.5	−53.5	−9.0
Pour Point	C	-	−93.1	-	−59.5	−12.3
Aniline Point	C	-	49.3	-	57.1	70.6
Smoke Point	mm	-	-	-	24.7	19.6
Liminometer		-	-	-	55.3	41.9
Cetane Index		-	35.1	-	44.1	53.1
Heat Value	kcal/kg	10578	10469	10542	10396	10234
RONC		56.27	39.53	46.37	18.56	-
RON + 0.15 g Pb/l		71.69	55.15	61.78	37.02	-
MONC		58.21	41.21	47.80	19.07	-
MON + 0.15 g Pb/l		69.95	54.82	61.40	34.83	-
Nitrogen	wt%	0.0	0.00	0.00	0.00	0.00
Vanadium	ppm	-	-	-	-	-
Nickel	ppm	-	-	-	-	-
RI at 70°C		1.3974	1.3974	-	1.4165	1.4470
Concarbon	wt%	-	-	-	-	-
Asphaltenes	wt%	-	-	-	-	-
BMCI		7.55	17.73	14.28	20.89	27.78
	wt%	**lv%**		**wt%**	**lv%**	
Cyclopentane	0.09	0.09	Total CC5			Total t
C6 iso paraffins	1.21	1.53				
C6 n-paraffins	1.58	1.99				
C6 naphthenes	0.49	0.53				
Benzene	0.09	0.08				

			Total C6s	3.36	4.13	Total t
C7 iso paraffins	1.18	1.44				
C7 n-paraffins	1.55	1.89				
C7 naphthenes	0.71	0.78				
Toluene	0.47	0.45				
			Total C7s	3.91	4.55	Total t
C8 iso paraffins	1.39	1.64				
C8 n-paraffins	1.41	1.67				
C8 naphthenes	0.80	0.86				
C8 aromatics	0.71	0.68				
			Total C8s	4.31	4.86	Total t
C9 iso paraffins	1.30	1.51				
C9 n-paraffins	2.55	2.96				
C9 naphthenes	0.76	0.80				
C9 aromatics	0.44	0.42				
			Total C9s	5.05	5.70	Total t
C10 iso paraffins	1.37	1.54				
C10 n-paraffins	1.21	1.38				
C10 naphthenes	0.78	0.80				
C10 aromatics	0.88	0.84				
			Total C10s	4.24	4.56	Total t

C11 iso paraffins	1.25	1.39				
C11 n-paraffins	1.06	1.19				
C11 naphthenes	1.06	1.08				
C11 aromatics	0.86	0.81				
			Total C11s	4.23	4.47	Total t
Crude sulfur	% wt	1.320	Crude Sg	Sg	0.8410	Crude
Crude viscosity 20	cst	5.990	Crude viscosity 30	cst	4.8100	Crude viscosi
Crude Pour Point	C	−39.000	Crude Acidity	mg koh/g	0.0500	Crude Conca
Crude Asphaltenes	%wt	0.3500	Crude Vanadium	ppm	4.0000	Crude
Crude salt content	lb/1000bb	2.1000	Crude water content	% vol	0.0500	N-para C5-95
N-paraffins 95-175	% wt.	30.6000	N-paraffins C5-149	% wt	40.3000	Acidit 232
Hydrogen 149-232	% wt	14.0500	Naphthalenes 149-232	% vol	0.03200	Color 149-23
Wax content 232-342	% wt	2.2000	Acidity 232-342	mg koh/g	0.01000	Color 232-34
Wax content 342-369	% wt	14.9000	Wax content 369-509	% wt	12.300	Basic 509
Organic O_2 369-509	% wt	0.0500	Acidity 369-509	mg koh/g	0.0200	Wax c 509-55
Basic N_2 509-550	ppm	225.0000	Acidity 509-550	mg koh/g	0.0600	Wax c 342+
Wax content 369+	% wt	11.0000	Wax Melt Pt. 369+	C	53.000	Acidit
Xylene Equiv. 369+		12.5000	Resins A 369+	% wt	8.000	Resins

Pen At 25c 509+	m	0.0500	Softening Point 509+	C	25.0000	Wax c 509+
Wax Melt Pt. 509+	C	62.0000	Pen At 25c 550+	m	0.0500	Soften Point 5
Xylene Equiv 550+		10.0000	Resins A 550+	% wt	12.0000	Resins
Mcpentene 15-95	% wt	0.2600	Hexane 15-95	% wt	0.19000	C6 – is 15-95
22-dmthlbutane 15-95	% wt	0.0100	23-dmthlbutane 15-95	% wt	0.0800	2-mthl 15-95
3-mthlpentane 15-95	% wt	0.4800	N-heptane 15-95	% wt	0.66000	Mthyl 15-95

Table 11-4 Middle East: Arabian Heav

Crude Data: Sample Date: SG:	**01/08/93 0.88874**	**API Gravity:**	**28.0**	**Analysis Date: Sulfur:**	**2.90 wt%**	
Light Hydrocarbon Analysis						
	wt %	**lv%**				
Hydrogen sulfide	0.00	0.00				
Methane	0.00	0.00				
Ethane	0.00	0.00				
Propane	0.07	0.12				Total to C3
Isobutane	0.11	0.16				
N-butane	0.77	1.11	Total C4s	0.88	1.27	Total to C4
Isopentane	0.70	0.94				
N-pentane	1.40	1.86	Total C5s	2.10	2.80	Total to C5

Wide Cut Data TBP Cut Points	C	C5–95	95–175	C5–149	149–232	332–342
Yield on Crude	wt%	5.62	9.12	11.21	10.90	15.90
	lv%	7.46	10.85	14.21	12.34	16.69
SG		0.6683	0.7456	0.6997	0.7839	0.8396
API Gravity		80.22	58.29	70.73	49.02	37.04
Watson K factor		12.55	12.12	12.31	12.00	11.94
Sulfur	wt%	0.002	0.024	0.006	0.153	1.379
Mercaptan Sulfur	wt%	0.001	0.001	0.001	0.001	0.001
Iso Paraffins	wt%	39.38	32.51	35.86	31.30	-
Total Paraffins	wt%	88.63	66.79	79.14	57.98	-
Naphthenes	wt%	10.58	22.37	16.55	23.06	-
Aromatics	wt%	0.79	10.84	4.31	18.96	30.07
Viscosity at 20°C	cSt	-	0.87	-	1.40	5.31
30°C	cSt	-	0.77	-	1.22	4.11
50°C	cSt	-	0.62	-	0.95	2.69
80°C	cSt	-	0.48	-	0.70	1.66
100°C	cSt	-	0.40	-	0.58	1.29
Cloud Point	C	-	−82.2	-	−58.6	−15.20
Freeze Point	C	-	−778.2	−98.3	−54.7	−11.3
Pour Point	C	-	−82.2	-	−58.6	−15.3
Aniline Point	C	-	54.8	-	59.7	68.0
Smoke Point	mm	-	-	-	26.1	-
Liminometer		-	-	-	58.7	-
Cetane Index		-	38.7	-	45.6	51.5

Heat Value	kcal/kg	10591	10477	10552	10401	10214
RONC		57.35	36.81	46.42	14.85	-
RON + 0.15 g Pb/l		72.51	52.33	61.62	33.56	-
MONC		59.45	38.26	47.92	15.12	-
MON + 0.15 g Pb/l		70.87	51.91	61.51	31.15	-
Nitrogen	wt%	0.0	0.00	0.00	0.00	0.00
Vanadium	ppm	-	-	-	-	-
Nickel	ppm	-	-	-	-	-
RI at 70°C		-	1.424	-	1.4362	1.4596
Concarbon	wt%	-	-	-	-	-
Asphaltenes	wt%	-	-	-	-	-
BMCI		8.23	15.00	11.52	19.74	30.31
	wt%	**lv%**		**wt%**	**lv%**	
Cyclopentane	0.05	0.06	Total CC5			Total to CC5
C6 iso paraffins	0.86	1.15				
C6 n-paraffins	1.05	1.41				
C6 naphthenes	0.31	0.36				
Benzene	0.03	0.03				
			Total C6s	2.25	2.94	Total to C6
C7 iso paraffins	0.80	1.03				
C7 n-paraffins	1.02	1.31				
C7 naphthenes	0.49	0.57				
Toluene	0.15	0.16				
			Total C7s	2.46	3.06	Total to C7
C8 iso						

C8 iso paraffins	0.89	1.11				
C8 n-paraffins	0.88	1.10				
C8 naphthenes	0.53	0.61				
C8 aromatics	0.38	0.39				
			Total C8s	2.68	3.21	Total to C8
C9 iso paraffins	0.89	1.09				
C9 n-paraffins	1.71	2.10				
C9 naphthenes	0.57	0.64				
C9 aromatics	0.22	0.22				
			Total C9s	3.39	4.05	Total to C9
C10 iso paraffins	0.91	1.09				
C10 n-paraffins	0.77	0.93				
C10 naphthenes	0.61	0.66				
C10 aromatics	0.48	0.48				
			Total C10s	4.24	3.15	Total to C10
C11 iso paraffins	0.84	0.99				
C11 n-paraffins	0.67	0.80				
C11 naphthenes	0.90	0.97				
C11 aromatics	0.50	0.50				
			Total C11s	2.92	3.26	Total to C11
Crude sulfur	% wt	2.900	Crude Sg	Sg	0.8905	Crude API

Crude sulfur	% wt	2.900	Crude Sg	Sg	0.8905	Crude API
Crude viscosity 20	cst	47.100	Crude viscosity 30	cst	30.800	Crude viscosity 40
Crude Pour Point	C	−51.000	Crude Acidity	mg koh/g	0.2300	Crude Concarbon
Crude Asphaltenes	%wt	4.500	Crude Vanadium	ppm	60.000	Crude Nicke
Crude salt content	lb/1000bb	2.6000	Crude water content	% vol	0.0500	N-paraffins C5-95
N-paraffins 95-175	% wt.	32.2000	N-paraffins C5-149	% wt	41.6000	Acidity 149-232
Hydrogen 149-232	% wt	14.1400	Naphthalenes 149-232	% vol	0.4500	Naphthalene 232-342
Color Stab 149-232		1.000	Wax content 232-342	% wt	2.9000	Acidity 232-342
Color Stab 232-342		1.000	Wax content 369-509	% wt	13.3000	Wax content 369-509
Basic N2 369-509	ppm	165.000	Organic O2 369-509	% wt	0.0500	Acidity 369-509
Wax content 509-550	% wt	9.10000	Basic N2 509-550	ppm.	365.000	Acidity 509-550
Wax content 342+	% wt	6.500	Wax content 369+	% wt	6.000	Wax Melt Pt 369+
Acidity 369+	mg koh/g	0.2300	Xylene Equiv. 369+		17.500	Resins A 36
Resins B 369+	% wt	9.000	Pen At 25c 509+	m	0.0076	Softening Point 509+
Wax content 509+	% wt	3.5000	Pen At 25c 550+	m	0.0022	Softening Point 550+
Xylene Equiv 550+		17.500	Resins A 550+	% wt	18.000	Resins B 55
Mcpentene 15-95	% wt	0.1800	Chexane 15-95	% wt	0.1100	C6 – iso-par 15-95
22-dmthlbutane	% wt	0.0100	23-dmthlbutane	% wt	0.0700	2-mthlpenta 15-95

15-95			15-95			15-95
3-mthlpentane 15-95	% wt	0.3900	N-heptane 15-95	% wt	3900	Mthylcyhex 15-95

11.4. Crude Oil Blending

Many refineries process more than one crude oil each being transported to the refinery via pipelines, ships or trucks. Since most refineries do not have adequate crude oil storage to isolate the various crude oils the storage tanks contain a mixture of the crude oils received by the refinery. Generally, crude blending is achieved by mixing good quality crudes with heavy low cost crudes to improve overall profitability for the refinery.

11.5. Laboratory Testing of Crude Oils

11.5.1. True Boiling Point (TBP) Curve

The composition of any crude oil sample is approximated by a true boiling point (TBP) curve. The method for determining this is a batch distillation operation using a large number of stages, usually >60, and high reflux to distillate ratio (>5). The temperature at any point on the temperature-volumetric yield curve represents the true boiling point of the hydrocarbon material present at the given volume per cent point distilled. TBP distillation curves are generally found only for the crude and not for petroleum products. Figure 11-3 shows a typical TBP curve.

Figure 11-3

Atmospheric TBP distillation curve

(Source: Kaes, G. L., Refinery Process Modeling: A Practical Guide to Steady State Modeling of Petroleum Processes, Kaes Enterprises, Inc., 2000).

The volatility of crude oil and petroleum fractions is characterized in terms of one or more of the laboratory distillation tests that are summarized in Table 11-5, namely the American Standard Testing Materials (ASTM) D86 and D1160 tests. The ASTM D86 distillation is a rapid batch distillation involving the equivalent of approximately one equilibrium stage and no reflux except for that caused by heat losses. The temperatures are reported at a pressure base of 1 atmosphere (760 mm Hg). For high altitude laboratories the barometric pressure corrections to the observed temperatures are significant and must not be ignored. Table 11-6 shows a sample of these temperature corrections.

Table 11-5 Laboratory Distillation Tests

Source; Perry's Chemical Engineer's Handbook, 7th ed., McGraw-Hill, 1997.

Test name	Reference	Main applicability
ASTM (atmosphere)	ASTM D 86	Petroleum fractions or products, including gasoline turbine fuels, napthas, kerosines, gas oils, distillate fuel oils, and solvents that do not tend to decompose when vaporized at 760 mm Hg
ASTM [vacuum often 10 torr (1.3 kPa)]	ASTM D 1160	Heavy petroleum fractions or products that tend to decompose in the ASTM D86 but can be partially or completely vaporized at a maximum liquid temperature of 750°F (400°C) at pressures down to 1 torr (0.13 kPa)
TBP [atmospheric or 10 torr (1.3 kPa)]	Nelson*, ASTM D 2892	Crude oil and petroleum fractions
Simulated TBP (gas chromatography)	ASTM D 2887	Crude oil and petroleum fractions
EFV (atmospheric		

(atmospheric, superatmospheric, or subatmospheric)	Nelson[‡]	Crude oil and petroleum fractions
*Nelson, Petroleum Refining Engineering, 4th ed. Mc. Graw-Hill, New York, 1958, pp 95–99.		
†Ibid, pp 104–105.		

Table 11-6 Temperature Corrections for Barometric Pressure

Source: Gerald L. Kees, Refinery Process Modeling: A Practical Guide to Steady State Modeling of Petroleum Processes, The Athens Printing Co., Athens, Georgia, 2000.

Laboratory pressure, mmHg							
Observed temp, °F	**572**	**597**	**622**	**648**	**673**	**698**	**724**
50	12	10	8	7	5	4	2
100	13	11	9	8	6	4	2
150	14	12	10	8	6	4	3
200	15	13	11	9	7	5	3
250	16	14	12	10	7	5	3
300	17	15	13	10	8	6	3
350	18	16	13	11	8	6	4
400	19	17	14	12	9	6	4
450	21	18	15	12	9	7	4
500	22	19	16	13	10	7	4
550	23	20	17	14	11	7	4
600	24	21	18	14	11	8	5
650	25	22	18	15	12	8	5
700	26	23	19	16	12	9	5
For use with ASTM D86 Distillation							
Temperature corrections are in °F.							

11.5.2. ASTM D86 Distillation

D86 distillation uses a simple Engler flask containing a calibrated thermometer to measure the temperature of the vapor at the inlet to the condensing tube. An inclined brass condenser in a cooling brass is attached, to remove all distilled vapors; no liquid reflux is returned to the flask. The condenser tube is cooled in an ice water bath to maintain the condensing temperature between 32 and 40°F (0–4.5°C). The light components that boil at temperatures lower than this are lost from the distilled product.

100 ml of sample is distilled and the vapor temperature against volume recovered is recorded. The *initial boiling point* (IBP) is defined as the temperature at which the first drop of liquid leaves the condenser tube. The *final boiling point* (FBP) or "end point" is the highest temperature recorded during the test. The total volume of the distillate is recorded as the recovery. Any liquid left in the still after the end point temperature is recorded is cooled and measured as the residue. The difference between 100 ml (initial sample volume) and the sum of the recovery and the residue is referred to as the loss. Repeated tests give for the initial boiling and end points. Intermediate distillation points are reproducible within 2 ml of distillate which corresponds to 6–7°F (3.3–3.9°C). Figure 11-4 shows a modern TBP apparatus.

Figure 11-4

The True Boiling Point (TBP) distillation apparatus.

11.5.3. Boiling Points

Figure 11-5 illustrates the interconversion between ASTM D-86 distillation 10% to 90% slope, and the different boiling points used to characterize fractions of crude oil. This is done to determine properties such as the volumetric average boiling (VARP) point, weight average boiling point (WABP), molal average boiling point (MABP), mean average boiling point (MeABP) and cubic average boiling point (CABP). On the basis of ASTM D-86 distillation data, the VARP is [290]:

(11-6)

where the subscripts 10, 30, 50, 70 and 90 refer to the volume per cent recovered during the distillation. The 10% to 90% slope used as the abscissa in Figure 11-5 is:

(11-7)

Figure 11-5

Characterizing Boiling Points of Petroleum Fractions

(From API Technical Data Book). Used by permission, Gas Processors Suppliers Association Book Data, 12th Ed., v.1 and 2., (2004).

The use of Figure 11-5 is as follows:

- to locate the curve for the distillation VARP in the appropriate

set for the type of boiling point desired.

• for a known 10–90% slope, to read a correction for the VABP from the selected VABP curve.

Example 11-1

Determine the mean average boiling point (MeABP) and the molecular weight for a 56.8° API petroleum fraction with the following ASTM distillation data:

(Source: Engineering Data Book: Gas Processors Suppliers Association, Tulsa, Oklahoma, 12th Ed., 2004)

% Over	Temperature, °F
IBP	100
5	130
10	153
20	191
30	217
40	244
50	280
60	319
70	384
80	464
90	592
EP	640

Solution

IBP = initial boiling EP = end point

Slope (592 – 153)/80 = 5.49

VABP = (153 + 217 + 280 + 384 + 592)/5 = 325°F

From Figure 11-5, Read down from a slope of 5.49 on the x axis to the interpolated curve closest to 325°F, in the set drawn with the dashed lines (MeABP). Read across to obtain a correction value of –54 on the ordinate.

MeABP = 325 – 54 = 271°F.

11.5.4. ASTM D1160 Distillation

This is designed for high boiling point samples in which there is significant cracking at atmospheric pressure. Typical samples for which D1160 distillations are useful are heavy heating oils, cracker gas oil feeds, and residual oils. The test is similar to the ASTM 86 method, but the sample is distilled at a reduced pressure to inhibit cracking. The distillation pressure commonly selected for ASTM D1160 tests is 10 mm Hg. All observed D1160 temperatures are pressure corrected to their equivalents at 760 mm Hg.

At 10 mm Hg, petroleum stocks can be distilled to a maximum boiling point temperature equivalent to 950–1000°F (510–538°C) at 760 mm Hg. Lower pressures are sometimes used to extend this boiling point range for residual stocks. The reduced pressure used for D1160 distillations produces a better separation of components than D86 distillation. The D1160 method closely approximates the TBP method, and the temperatures recorded for 50% distilled and higher samples are assumed to be identical to the corresponding TBP temperatures. The D1160 method is sometimes used to extend the TBP distillation curve for mixtures containing high boiling materials.

11.5.5. Determination of ASTM IBP, 10%, 20–90% Points of Blend

This method is applicable to blends containing distillate stocks having an ASTM initial boiling point higher than 85°F, and an ASTM end point lower than 700°F. It is based on the observation that a straight summation line can be drawn through an ASTM distillation point of a

blend.

The slope of the line must be such that the sum of the proportions of each blend component corresponds to its intersection with the ASTM distillation curve. For TBP distillation, the summation lines are parallel to volume per cent axis on an ASTM distillation plot. ASTM summation lines slope due to poor fractionation of ASTM distillation and the slope varies according to distillation end point. The following are the slopes for varying distillation points [299]:

Distillation Point	**Slope of Summation line, °F**
IBP	–180°F per 100% distilled
10%	–180°F per 100% distilled
20%	–100°F per 100% distilled
30%	–80°F per 100% distilled
40%	–50°F per 70% distilled
50–90%	–20°F per 70% distilled

11.5.6. ASTM 10-90% Points

ASTM distillation curves are drawn for each blend component with the temperature on the vertical axis and the volume per cent distilled on the horizontal axis. Distillation must be done on a consistent basis for all components, *i.e.* either percent evaporated or percent recovered.

A guess is made of the temperature at which a given proportion of the blend is distilled, and the corresponding point is marked on the graph. A summation line of specified slope is drawn through the point. The vol% distilled is read off vertically below the intercept of the summation line and ASTM distillation curve of each component (Figure 11-6). The sums for all blend components should equal the proportion of blend originally estimated, otherwise a new guess of

temperature at which the specified proportion of blend is distilled is made and the procedure repeated.

Figure 11-6

ASTM distillation blending procedure

(source: Parkash, R., Refining Processes Handbook, Elsevier: Gulf Professional Publishing, 2003).

Where the second estimate does not produce the required result, an interpolation is made between the two estimates.

11.5.7. Initial Boiling Point Determination

This method is identical to 10–90% points of blend, except that the distillation curves for the components are extrapolated to –1.4%. Therefore, –1.4% becomes the new zero of the modified scale, and 10% becomes 11.4%. The volume distilled at 1.4% is next estimated by the previous procedure to give the IBP of the ASTM curve.

11.5.8. ASTM End Point of Blend

The end point of a two-component blend is referred to as the function of the end point, proportion of blend component, and slope of the tail of the distillation curve of the higher-boiling component.

11.5.9. Flash Point

The flash point is a measure of the volatile components in a petroleum middle distillate. It is dependent on the predicted solubility of light hydrocarbons in the middle distillate. Various attempts have been made to correlate flash points for petroleum distillates, and all attempts use the predicted ASTM 86 10% distilled temperature as a correlating parameter.

The flash point test determines the temperature at which the vapors from petroleum oil will ignite or “flash”. Three basic procedures are used to determine this: the Cleveland open cup method (ASTM D92), the Penskey Martens (ASTM D93), and the Tag (ASTM D53) closed cup methods.

A sample is heated at a specified rate and an open flame introduced into the emitted vapors at specified time intervals. The temperature at which the vapors ignite is recorded as the flash point. The open cup method is often used because of its simplicity. Below flash points of 510°F (266°C), the open cup flash point is approximately 10°F (5.6°C) higher than the closed cup flash point for the same sample. The open cup method heats the oil at a rate of 10°F (5.6°C) per minute with the test flame applied at 30 sec. intervals. Accuracy for this method is ±5°F (±2.8°C). The closed cup tester heats the oil at a rate of 1.8°F per minute and is the more precise of the two methods.

11.5.10. Flash Point, °F, As a Function of Average Boiling Point

The correlation is applicable up to 1050°F [299]:

Flash =	0.535T – 93, if (T ≤ 174°F)
Flash =	0.67T – 116, if (174 ≤ T ≤ 303°F)
Flash =	0.7506T – 140.3, if (303 ≤ T ≤ 577°F)
Flash =	0.674T – 94.5, if (577 ≤ T < 1050°F)

11.5.11. Smoke Point of Kerosenes

The smoke point (S) of kerosene can be estimated by the following correlation with the aromatic content of the stream.

(11-8)

where

S = Smoke point, mm
arom = aromatic content, LV %
API = API gravity of the cut.

11.5.12. Luminometer Number

The luminometer number is qualitatively related to the potential radiant heat transfer from the combustion products of aviation turbine fuels. It provides an indication of the relative amounts of radiation emitted by combustion products of gas turbines fuels.

The luminometer number (L_n) is related to the smoke point by the following correlation (ASTM specification D-1740):

(11-9)

(11-10)

where
L_n = the luminometer number
SP = the smoke point

11.5.13. Reid Vapor Pressure (RVP)

The Reid vapor pressure test is used to determine the front-end volatility of products in the gasoline through heavy reforming naphtha boiling point range. It correlates with the normal butane content of the sample and the RVP of gasoline blends is adjusted by adding or removing normal butane. For gasoline, the RVP is 8–14 psi, and the percentage of butanes in the stream is the same as the RVP. The value is always reported in absolute pressure units of psi, kPa, bar, *etc*.

The test procedure for gasolines and naphthas involves combining one part of chilled liquid sample at 36°F (2.6°C) with four parts of laboratory air at 70°F (21.4°C) in a bomb that has a pressure gauge attached to the lid. The bomb is placed in a constant temperature bath at 100°F (37.8°C), and when the pressure reaches a constant value, this is recorded as the Reid vapor pressure. Table 11-7 shows the Reid vapor pressure for gasoline streams with the largest deviations occurring for the samples at the highest vapor pressures. The Reid vapor pressure measurements have been used to assess the amount of light hydrocarbons that may be lost during filling and draining operations of tanks [296].

(11-11)

Table 11-7 Relationship between the True and Reid Vapor Pressure for Miscellaneous Volatile Products

Source: W.L. Nelson, Petroleum Refinery Engineering, 4th. Ed., McGraw-Hill Book Co., New York, p 133, [1958].

	Vapor pressure, psi		**Ratio**
Product	**Reid**	**True**	**True/Reid**
Natural gasoline	24.0	26.2	1.09
Natural gasoline	20.0	21.8	1.09
Natural gasoline	18.4	20.1	1.092
Natural gasoline	17.8	20.3	1.14
Natural gasoline	16.0	17.5	1.09
Light gasoline	14.0	15.4	1.10
Light gasoline	12.0	12.9	1.07
Natural gasoline	12.0	12.9	1.07
Gasoline	10.0	10.4	1.04
Straight-run gasoline	8.3	9.7	1.17

Gasoline	8.0	8.3	1.04
Gasoline	6.0	6.3	1.04
Refinery gasoline	5.9	7.2	1.16
Gasoline	5.0	5.2	1.04
Gasoline	3.0	3.1	1.03
Gasoline	2.0	2.1	1.05

The Reid vapor pressure is expressed in psi (1 psi = 6.9 kPa). Table 11-8 shows the RVP of some crude oils. Measurement of the vapor pressure and flash point of crude oils indicates the amount of light hydrocarbons present. Crude oils with a vapor pressure of >0.2 bar at 37.8°C have a flash point <20°C.

Table 11-8 RVP and Flash point of some crude oils

Country	Crude	RVP, bar	Flash point, °C
Iraq	Kirkuk	0.29	<20
Kuwait	Kuwait	0.51	<20
Nigeria	Nigerian	0.26	<20
Qatar	Qatar	0.50	<20
Venezuela	Bachaquero	0.06	46

11.5.14. Vapor Pressure of Narrow Hydrocarbon Cuts

Vapor pressure of narrow hydrocarbon cuts can be estimated from their boiling points by the correlation of Van Kranen and Van Ness [299]:

(11-12)

where

(11-13)

(11-14)

where

B = average boiling point, °C

T = temperature, °C

VP_{mmHg} = vapor pressure, mmHg

VP_{psi} = vapor pressure, psi

11.6. Viscosity

The viscosity of oil is its internal resistance to flow and a measure of its usefulness as a lubricating substance. Crude oil viscosities are usually used for calculating the pressure drop in pipelines, and for deriving the specifications of equipment such as exchangers and pumps. The Saybolt Universal viscosity test (ASTM D88) is the most common method used by refiners. This measures the time taken (in seconds) for a sample of 60 cc of oil to flow from a container through a calibrated orifice at a prescribed temperature. The usual test temperatures are 100 and 210°F (37.8 and 99.2°C).

Saybolt viscosities are referred to as seconds. The Saybolt Furol viscosity test (ASM D88) is a similar test for heavy fuel oils and uses a larger orifice. Saybolt Furol viscosities are reported at 122 and 210°F (50.3 and 99.2°C).

11.6.1. Conversion to Saybolt Universal Viscosity

Kinematic viscosity in centistokes (mm^2/s) can be converted to Saybolt universal viscosity in Saybolt universal seconds (SUS) units

at the same temperature by API databook [300] procedure or using ASTM conversion tables [301].

(11-15)

where

SUS = Saybolt universal viscosity in Saybolt universal seconds units at 37.8°C

cSt = kinematic viscosity, centistokes (mm^2/s)

where

SUS = SUS viscosity at 37.8°C

SUS_t = SUS viscosity at the required temperature

t = temperature at which SUS viscosity is required, °C

11.6.2. Conversion to Saybolt Furol Viscosity

Kinematic viscosity in centistokes (mm^2/s) at 50 and 98.9°C can be converted to Saybolt Furol viscosity in Saybolt Furol seconds (SFS) units at the same temperature by API databooks [302]:

(11-16)

(11-17)

where

SFS_{50} = Saybolt Furol viscosity at 50°C, Saybolt Furol seconds

VIS_{50} = kinematic viscosity at 50°C, centistokes (mm^2/s)

$SFS_{98.9}$ = Saybolt Furol viscosity at 98.9°C, Saybolt Furol

seconds

$VIS_{98.9}$ = kinematic viscosity at 98.9°C, centistokes (mm^2/s)

11.6.3. Equivalents of Kinematic (cSt), Saybolt Universal (SUS), and Dynamic Viscosity

Saybolt Universal Seconds = Centistokes × 4.6347 for

Saybolt Furol Seconds = Centistokes × 0.4717

11.7. Octanes

The octane test is used to measure the knocking characteristics of a fuel in a laboratory gasoline engine. The engine is calibrated by measuring the knocking value of a standard material; 2-2-4 trimethyl pentane (224 TMP). By definition, an octane number of 100 is assigned to 224 TMP. The knocking value of the test sample is compared to that of the standard member to determine its octane member.

11.8. Cetanes

The cetane number is used to measure the performance of a fuel in a diesel engine. Although determined in a laboratory engine (ASTM engine test D613), it is calculated with a formula that relates cetane number, API gravity and ASTM D86, 50% distilled.

All the laboratory tests are used to determine the qualities of feed stocks and products from petroleum processes. They are essential in design and simulation since they can be used to define petroleum pseudo-components to represent the streams for the unit operations calculations.

11.8.1. Cetane Index

The cetane index (CI) is estimated from the API gravity and 50% ASTM distillation temperature in °F of the diesel. It is defined by [303]:

(11-18)

or

(11-19)

where

API = API gravity
M = mid-boiling temperature, °F
D = density at 15°C
B = mid-boiling temperature, °C

11.9. Diesel Index

The diesel index (DI) is correlated with the aniline point of diesel as:

(11-20)

The relationship between diesel index and cetane index of diesel is:

(11-21)

11.10. Determination of the Lower Heating Value of Petroleum

Fractions

If the API gravity of a distillate is above 42°, the net heat (H) of combustion in Btu/lb can be determined as follows:

(11-22)

If the API gravity of distillate or residue is below 42°, the net heat (H) of combustion in Btu/lb can be determined as follows:

(11-23)

Correct the preceding heating values for sulfur content of the distillate or residue.

(11-24)

where

Heat = lower heating value, Btu/lb
K = Watson K
API = API gravity
S = Sulfur content (%)

11.11. Aniline Point Blending

The aniline point of a gas oil is indicative of its aromatic content. Aromatic hydrocarbons exhibit the lowest and paraffins the highest values. Aniline point (AP) blending is not linear, and therefore blending indices are used.

The conversion of aniline point to aniline index is given by:

(11-25)

or, alternatively the aniline point is:

(11-26)

where

ANLIND = the aniline index

AP = the aniline point

Nelson [304] presents four fundamental principles for the operation of crude columns and other multidraw columns as:

1. The yield of a given product is primarily a function of the composition of the feed mixture, not the degree of separation.
2. The number of trays only slightly alters the boiling range of the products as defined by ASTM initial boiling points (IBP) and end points.
3. The initial boiling point of side draw products is always low, and it must be corrected by either steam stripping or reprocessing.
4. The end point of a side draw product is primarily controlled by opening or closing the draw valve to change the yield.

11.12. Chromatographically Simulated Distillations

Gas chromatography methods have been developed to predict the boiling point distribution corresponding to that determined by the detailed TBP method. Four standard tests are currently employed by refinery test laboratories, namely; D3710, D2887, D2887 extended, and HTSD. Table 11-9 summarizes simulated distillation procedures. The methods of operation for each are similar, and the most

commonly used method is illustrated in ASTM Standard D 2887-73. This method is limited to petroleum fractions with a final boiling point of 1000°F (538°C). The sample must also have a boiling point range ≥100°F (55°C). Carbon disulfide (CS_2) is used for the carrier gas. The chromatographic column is first calibrated with a standard mixture of normal paraffin hydrocarbons. A calibration curve of retention time in the column versus normal boiling point is developed for normal paraffins by increasing the column temperature and recording the areas of each recorded peak of the individual components at timed intervals. The sample is next introduced into the column; the retention times and the areas of the sample components are recorded. The components in the sample are then assigned boiling points by using the calibrated curve developed for the normal paraffin standard.

Table 11-9 Simulated Distillation Procedures [15].

Method	Application	Standard	Range	Report basis
D3710	gasolines, naphthas	C_5–C_{15}	<500°F	volume %
D2887	cycle oils, gas oils	C_5–C_{44}	<1015°F	weight %
D2887 – ext.	heavy gas oil, crude	C_5–C_{60}	<1140°F	weight %
HTSD	heavy gas oil, crude	C_5–C_{120}	<1382°F	weight %
Note: The standard is a mixture of normal paraffins with carbon number ranges as shown.				

Crude oils and heavy residues have such high boiling points that the column temperature cannot be increased sufficiently to measure the highest boiling fractions. Hence, the total area of the sample cannot be known and the boiling point curve for the sample cannot be determined accurately. HTSD is a modification of the D2887 procedure and extends the boiling range of hydrocarbon to a final

boiling point of ≈1382°F (750°C). A careful choice of chromatographic conditions is used to cause the high boiling components to be eluted from the column some 500–600°F (260–316°C) below their normal boiling points, and as such provides great potential for analysis of heavy petroleum residues.

11.13. Process Description

Figure 11-7 shows the crude fractionation unit in more detail with the pumparound. Cold crude oil from storage is pumped through a series of exchangers, where it is preheated by exchange with intermediate products from the crude and vacuum distillation columns.

Figure 11-7

Atmospheric crude column with pumparound and pumparound reflux.

Crude desalting is the first processing step in a refinery. This is done to remove the salts, suspended solids and water from the unrefined crude before introducing it to the crude distillation unit (CDU). Salt is present in crude oil in the form of dissolved or suspended crystals in water that is emulsified with the crude oil. The desalting process washes the salt from crude oil with water. It can often be difficult to produce efficient and economical water/oil mixing, water wetting of suspended solids, and separation of oil from wash water. The separation of oil and wash water is affected by the specific gravity, viscosity and pH of the crude oil, and the ratio of water/crude used for washing.

The crude oil is injected with five to six liquid volume percent water at a temperature 200–300°F (93–149°C). Both the ratio of water to oil and the operating conditions are determined by the specific gravity of the crude oil, and typical operating conditions are shown in Table 11-10. The salts dissolve in the wash water, and then the oil and water phases are separated in a settling vessel, either by adding chemicals to break up the emulsion, or by the application of an electrostatic field to coalesce the droplets of salt water. Either an AC or DC field may be applied at potentials of 16,000–35,000 V. In the desalter, the salts (e.g. magnesium chlorides) selectively migrate to the aqueous phase which forms a brine solution at the bottom of the desalter, while the crude oil floats above and forms a separate stream. Efficiencies of up to 90–95% water removal are achieved in a single stage, and up to 99% in a two-stage desalting process. Figure 11-8 shows an electric desalting method and Table 11-11 shows typical operating conditions in the desalting unit. After desalting, the crude oil contains from 0.25–0.5 liquid volume percent water from solubility and emulsion effects.

Table 11-10 Washwater requirements of Desalters

Crude oil API	Washwater, Vol. %	Temperature, °F
API > 40	3–4	240–260
30 < API < 40	4–7	260–280
API < 30	7–10	280–300

Figure 11-8

Two-stage desalter.

Table 11.11 Operating conditions

Source: Parkash, S., Refining Processes Handbook, Elsevier: Gulf Professional Publishing, Co. 2003.

Parameter	Units	Value
Crude to desalter*	bpsd	98000
Water to desalter	gpm	145
Water to crude ratio	%	5
Demulsifier injection	ppmw	10–15
Pressure		
Crude to desalter	psig	125
mixing valve	psi	20
Temperature		
Crude to desalter	°F	270
Water to desalter	°F	265
Crude from desalter	°F	260
Analysis results		
Crude inlet salt	lb/1000 bbl	3.94
Crude inlet salt	ppmw	12.87
Crude outlet salt	ppmw	1.2
Outlet BS & W	% Mass	0.05
Water		
Inlet salt content	ppm	100
Outlet salt content	ppm	310
Inlet oil content	ppm	7
Outlet oil content	ppm	10
pH inlet		6.5
Outlet pH		6.5
Outlet pH after NH_3		
Injection		7

Injection		
Note: Basis 98000 bpsd crude.		
*30.4 API Crude.		

The pressure of the desalted crude oil is then raised in a booster pump, and it continues through a preheat train where its temperature is raised by exchange with hot intermediate streams from the crude and vacuum columns. Upon exit from the last preheat exchanger the oil is heated to the desired column inlet temperature in the crude furnace. Furnace outlet temperatures typically range from 630–725°F (333–385°C), depending on the crude oil composition.

The hot oil mixture from the furnace then flows through a transfer line into the flash zone of the main fractionation column. Here it mixes with steam and light hydrocarbons from the column stripping zone. Here, sufficient steam is used to strip the lightest hydrocarbons from the gaseous portion of the oil that rises to the top of the column. The stripping steam also lowers the partial pressure of these hydrocarbons in the column flash zone. This in turn lowers the temperature required to distill the overhead products from the column feed.

The hot liquid from the flash zone drops through the column stripping section, where it is stripped by steam injected into the bottom of the column. The hot vapors from the flash zone rise, contact the colder reflux flowing down the column which condenses out the side liquid products. Reflux flow is provided by the overhead condenser and by pumparound circuits in which liquid drawn from the side of the condenser is pumped through exchangers and cooled by exchange with crude oil in the preheat train. The cooled pumparound liquid is typically returned to the column a few trays above the draw tray. Pumparound flow rates are high, because heat is being exchanged with crude oil that is at a high flow rate.

The pumparound cooling circuits accomplish three objectives: firstly, they remove latent heat from the hot flash-zone vapors and help condense the side products; secondly, they improve the efficiency of the crude preheat train by allowing heat recovery at higher temperature levels than an overhead condenser alone could achieve. This decreases the work required from the crude furnace. Finally, they reduce the vapor flow rate through the column so reducing required size. Since crude column sizes are primarily determined by the vapor flow rates, pumparound cooling is an essential factor in their design. However, the pumparound circuits degrade the fractionation between the side products by reducing the internal liquid flows between side products.

The liquid side draw products are usually stripped in steam strippers or recoiled strippers to remove light materials and control their flash points, so the composition of the lightest portion of a liquid side product is controlled by a side stripper, not the main column. In the case of the gas oil side product, stripping is used to remove diesel boiling range materials from the cracking stocks for feeding to fluid catalytic crackers (FCCs).

The maximum boiling point of the liquid side products are controlled on the main column by their draw rates. For example, to increase the maximum boiling point of the kerosene product, it is necessary to decrease the flow of the diesel product (which has a higher boiling range) and increase the flow of the kerosene product. This adjustment allows heavier components to travel up the column to the kerosene draw tray, increasing the maximum boiling point of the kerosene product. There may also be a decrease in the lightest portion of the diesel product because of this adjustment, but this effect is small when compared to the effect of the diesel steam stripper. Table 11-12 shows the products of crude oil distillation, and Table 11-13 shows the approximate boiling ranges for crude oil fractions.

Table 11-12 Products of crude oil distillation

Source: Koch Baby Majoorana, Modern Petroleum Chemistry: An overview, Kannatheri Publications Kochi, 2003.

Fraction	Boiling range	Remarks
1. Gases (mainly methane to the extent of 65–90% together with some ethane, propane and butane)	<20°C	Resembles natural gas and is useful for fuel and chemicals. In refineries most of these gases are burned in furnace as fuel due to the costs involved in the recovery
2. Naphtha		
a. Light Naphtha (C_5–C_6 hydrocarbons)	70–140°C	Useful for both fuel and chemicals. But light naphtha is not preferred in gasoline because on catalytic reforming yields the carcinogenic compound benzene
b. Heavy Naphtha (C_7–C_9 hydrocarbons	140–200°C	Base for gasoline
3. Atmospheric gas oil		
a. Kerosene (contains C_9–C_{16} compounds).	175–275°C	Used as heating fuel
b. Diesel fuel (contains C_{15}–C_{25} compounds, mostly linear)	200–370°C	Fuel for diesel engines
4. Heavy fractions		
a. Lubricating oil (contains C_{25}–C_{40} compounds).	>370°C	As lubricating oil
b. Residual or heavy fuel oil.	_____	Useful as boiler fuel

Table 11-13 Approximate ASTM boiling point ranges for crude oil fractionation

	Boiling range

Fractions	Boiling range	
	°F	°C
Light naphtha	85–210	30–99
Heavy naphtha	190–400	88–204
Kerosene	340–520	171–271
Atmospheric gas oil	540–820	288–438
Vacuum gas oil	750–1050	399–566
Vacuum residue	1000+	538+

11.14. Process Variables in the Design of Crude Distillation Column

The following are important variables in the design of a crude oil distillation unit:

1. The nature of the crude: water content, metal content and heat stability. The latter limits the temperature to which the crude can be heated in the furnace without incipient cracking.
2. Flash zone operating conditions: flash zone temperature is limited by advent of cracking; flash zone pressure is set by fixing the reflux drum pressure and adding the value to the line and tower pressure drop.
3. Over-flash: this is the vaporization of crude over and above the crude overhead and sidestream products. Over flash is generally kept in the range of 3–6 LV% (LV = liquid volume). Over-flash prevents coking of wash section plates and carryover of coke to the bottom sidestream, and also ensures a better fractionation between the bottom sidestream and the tower bottom by providing reflux to plates between the lowest sidestream and the flash zone. A larger over-flash also consumes more energy, thus over-flash is kept to the minimum needed for the the quality requirement of the bottom side stream.
4. Steam stripping: the bottom stripping steam is used to recover light components from the bottom liquid. In the flash zone of an atmospheric distillation column, approximately 50–60% of the

crude oil input is vaporized. The unvaporized crude travels down to the stripping section of the columns, where four to six plates strip out any low boiling point distillates still contained in the remaining crude. The steam rate used is approximately 5–10 lbs/bbl of stripped product. The flash point of the stripped stream can be adjusted by varying the steam flow rate.

5. Fractionation: this is the difference between the 5% point on the ASTM curve of a heavy cut and the 95% point on the ASTM curve of a lighter cut of two adjacent side products. A positive difference is referred to as a gap and a negative difference is referred to as an overlap.

11.14.1. Process Design of a Crude Oil Distillation Column

The following steps in the design of a crude distillation column are:

- Prepare TBP distillation and equilibrium flash vaporization curves for the crude to be processed. Several methods are available for converting TBP data to EFV curves.
- Using crude assay data, construct TBP curves for all products except gas and reduced crude. These are then converted to ASTM and EFV curves by Edmister [305], Maxwell [306] or computer methods.
- Prepare material balances of the crude distillation column, on volume and weight basis, showing crude input and product output. In addition, plot the physical properties, such as cut range on TBP and LV% and vol% vs. SG, molecular weight, mean average boiling point, and enthalpy curves for the crude and its various products.
- Fractionating requirements: Ideal fractionation is the difference between the 5% and 95% points on ASTM distillation curves obtained from ideal TBP curves of adjacent heavier and lighter cuts. Having fixed the gaps as a design parameter, the difference between ideal gap and actual gap required is referred to as the

deviation.

The deviation or gap can be correlated with an F factor, which is the product of the number of plates between two adjacent side draws and the internal reflux ratio. Internal reflux is defined as the volume of hot reflux at 60°F below the drawoff plate of the lighter product, divided by the volume of liquid products at 60°F, except gas lighter than the adjacent heavier products. This implies that the reflux ratio and the number of plates are interchangeable for a given fractionation.

11.14.2. A Case Study: Process Description of Light Arabian Crude Using UniSim® Simulation Software

Honeywell UniSim® simulation software was used to simulate the flow of crude oil from the preflash unit to the crude distillation fractionators. The following illustrates the process:

After characterization (Figure 11-9), the crude oil passes through a preheat train at 10,000 barrel/day at 29°API and is fed into a preflash separator operating at 450°F (232°C) and 75 psia. The vapor from this separator bypasses the crude furnace and is remixed with the hot crude at 650°F (343°C) preflash liquids leaving the furnace. The combined stream is then fed to atmospheric fractionation column as shown in Figure 11-10. The column has a total condenser, three coupled side strippers and three pumparound circuits. Light gases and a naphtha product are produced overhead, a kerosene product from the first recoiled side stripper, a diesel product from the second steam-stripped side stripper, and an atmospheric gas oil (AGO) from the third side stripper. This last is further processed in the vacuum distillation unit. Figure 11-11 shows a snap shot of the process flow diagram of the crude distillation unit with the splitters, and Table 11-14 shows typical operating conditions of the atmospheric distillation column.

Figure 11-9

True Boiling Point Distillation – Arab Extra Light 36.7

Figure 11-10

Process flow diagram of the crude distillation unit.

Figure 11-11

A snap shot of crude distillation unit with pumparounds and side-strippers.

Table 11-14 Atmospheric Tower Operating Conditions

Operating parameter	Units	Value
Temperature	°F	
Transfer line		660
Flash zone		657

Flash zone		657
Tower top		359
Kerosene drawoff		469
Pumparaound draw off		548
Pumparaound return		345
Light diesel draw off		603
Tower bottom		648
Pressure	psig	
Reflux drum		2.0
Tower top		10.3
Flash zone		14.7
Reflux ratio, reflux/Liq.dist.		0.6
Stripping steam		
To atmospheric tower	lbs/bbl Resid.	5.5
To kerosene stripper	lbs/bbl Resid	5.9
To diesel stripper	lbs/bbl Resid	2.1
Atmospheric heater		
Process fluid conditions		
Temperature in	°F	453
Temperature out	°F	660
Pressure drop	psi	138
Tube skin temperature (avg.)	°F	735
Stack gas temperature	°F	725
Fractionation efficiency		
5%–95% ASTM distribution gap		
Atmospheric naphtha-kerosene		Gap + 10
Kerosene-light diesel		Gap – 36
Note: Basis 154000 bpsd Kuwait Crude Run.		

The two basic steps in the process simulation are:

1. Setup involves listing the components between C_1–C_4 light

ends, as well as the pseudo components that will be used to represent the C_5^+ and above portion of the crude oil. The oil characterization procedure as described earlier is used to convert the laboratory data into petroleum pseudo-components.

2. Steady state simulation modeling involves simulating the prefractionation train, which consists of a separator and heater. This determines the feed to the atmospheric fractionators, and includes the preflash separator, crude furnace and mixer which recombine the preflash vapor and furnace outlet streams.

Simulation of the atmospheric fractionator involves adding the column streams inlets to the flowsheet and carrying out the appropriate calculation.

Before doing this, the steam feeds and the energy requirements of the column must be defined. Three streams are fed to various locations in the column. The atmospheric tower is represented by 29 ideal stages (not including the condenser). The pressures of the overhead condenser, the top stage and the bottom of the column are 19.7 psia; 28.7 psia and 32.7 psia respectively. The condenser therefore experiences a pressure drop of $\Delta P = 9$ psi. The temperature estimates for the condenser, top stage, and bottom stage are 100°F (37.8°C); 250°F (121°C) and 600°F (316°C) respectively. Water is removed via a side water draw from the three phase condenser.

Solution of fractionation calculations involves a trial and error procedure. Temperature and flow profiles are assumed for all trays in the column and vapor–liquid equilibrium and heat balances are calculated for each tray. All calculation procedures keep the column pressures constant. The results from the assumed profiles are used for the next series of calculations. Computations are further complicated by the fact that the equilibrium K values for the components on each

tray are affected by the tray composition as well as the temperature and pressure. Further, all of the tray-to-tray calculations are coupled into a large set of equations, which describes the component mass balances and the heat balances for every tray, the vapor-liquid equilibrium relationships between the trays and the overall heat and mass balances for the entire operation. Parameters are added to define process conditions or specifications that must be met by the final solution. A converged solution is defined as a set of independent variables, *i.e.* flows, heat inputs, tray temperatures, etc., which produce a solution for all of the column equations within certain specified tolerances. Tolerances are prescribed for the component material balances, the tray heat balances, the tray equilibrium relationships (e.g. expressed as the tray liquid bubble point condition), the performance specifications and the tray equilibrium K values. The latter tolerance is required because the K values are dependent on composition. Therefore, it is essential that the K-values converge together with all the column equations. Several methods have been described in the literature to solve distillation problems rigourously. One of the most useful method for petroleum distillation columns is the *inside-out* method, which is described elsewhere [255].

11.14.3. Column Conventions

Column conventions are similar for all process simulators. All distillation calculation methods are derived for theoretical trays, *i.e.* trays in which perfect vapor-liquid equilibrium is achieved. The condenser and overhead accumulator drum are considered to be the first tray. A net liquid product from this tray is returned to the column as reflux. Liquid and vapor products are withdrawn from the reflux drum. Where only a liquid product is withdrawn these conditions are not met. For many columns, the liquid product is subcooled below its bubble point. The same convention is followed and the overhead system is still referred to as the first tray. The top tray of the column is considered to be the second tray for a column with a condenser. The

tray at the bottom is referred to as the Nth tray.

The reboiler is normally considered to be the highest tray number, labelled N + 1. Columns may have side liquid and vapor draws and side water draws. Side stripping columns (e.g. for kerosene, light and heavy gas oils, diesel, etc.) may also be attached to the column. Conditions for side columns are solved along with the main column, by the inside-out method. Pumparound side draws are used to remove heat from the column, which may have multiple feeds. UniSim designates the condenser-reflux drum trays as the *condenser* and the second tray as tray number one. It also designates the *reboiler* (when present) as the reboiler and tray N as the highest tray number in the column model. For columns with no condenser-reflux drum, *e.g.* a recoiled stripper, the top tray of the column is designated as tray one by all simulators. For columns with no reboiler, *e.g.* an absorber, the bottom tray of the column is designated as tray N by all simulators.

11.14.4. Performance Specifications Definition

Poorly chosen specifications are the cause of nearly all convergence failures when solving distillation problems using the inside-out method. Additionally, specifications for the column model should be similar to those used in actual operation so actual operating experience is useful for the modeler. For every fixed value imposed on a column model, there must be a corresponding degree of freedom or unspecified parameter. Changing these parameters enables the user to find sets of conditions for the column that meet the specified performance criteria. For distillation models, the choices of parameter to vary are the column heat sinks, the side product draw rates, and the column heat sources. For example, a conventional column with two products, a reboiler and a condenser has two possible degrees of freedom and so may have two performance criteria specified.

11.14.5. Troubleshooting Actual Columns

Troubleshooting with computer simulation must be dependent on mathematically converged column solutions, as it is impossible to attain valid conclusions using simulation models for which the column mathematical constraints are unsatisfied. Thus the first stage in troubleshooting is to ensure converged results with the column model. The column models must be ascertained as being in predictive mode for troubleshooting. The models based on local tray efficiency models will yield poor results because they cannot predict an alternate operating point. Computer simulation has successfully being applied to petroleum distillation problems; however the necessity of representing petroleum mixtures with pseudo-components has been a serious impediment to the development of accurate models. Therefore, where a computer model matches the actual operation, there are definitive reasons. Differences between the model and the plant are commonly reflected in the product properties, product compositions, product rates, column temperatures, top reflux rates, internal liquid flow rates and heat duties. Kaes [307] has described various step wise approach when matching models to plant data.

Figure 11-12 and Figure 11-13 show temperature and pressure versus tray position from the top. Zones of relatively constant temperature in the column indicate tray damage or missing trays.

Figure 11-12

Temperature vs. Tray Position from Top.

Figure 11-13

Pressure vs. Tray Position from Top.

Simulation program of the crude oil using UniSim® software (**crude-oil-distillation.usc**) can be accessed from the Elsevier website. Results of the simulation of crude oil distillation are shown in Appendix L, and Table 11-15 shows rules for the design and operation of petroleum fractionators.

Table 11-15 Some Rules for Design and Operation of Petroleum Fractionators

Oil refineries are often designed to handle various crude oils with varying boiling assays. The refinery responds to market conditions by making different product ranges at different times of the year. The crude oil fractionation column and associated equipment (e.g. heat exchanger networks) must be flexible enough to handle these changes, whilst maintaining the boiling point curves for every product. This waves designing a fractionating column an even more complex operation.

11.15. Gas Processing

Gas processing or treating may be defined as removal of hydrogen sulfide (H_2S) and carbon dioxide (CO_2) from a feedstock. H_2S must be removed for health reasons and to prevent corrosion. Natural gas pipeline specifications require no more than ¼ grain/100 SCF. This is

equivalent to 4 ppmv or 7 ppmw (for a 0.65 specific gravity gas). By contrast, the human nose can detect 0.13 ppmv and the threshold limit value for prolonged exposure is 10 ppmv.

CO_2 must be removed to prevent corrosion and because it lowers the heating value of natural gas. Some processes allow most of CO_2 to "slip" while selectively removing H_2S. Where CO_2 can be tolerated downstream, it is more economical to remove only H_2S. H_2S can be selectively removed in order to enrich the H_2S content of the acid gas stream feeding a sulfur recovery unit. However, in such cases the concentration of CO_2 should be retained below 80% in the acid gas feed. Some of the reasons for selective H_2S removal are:

- H_2S enrichment for sulfur recovery
- Economical design (capital and operating costs)
- Special cases such as energy recovery.

11.15.1. Natural Gas

Natural gas is mixture of hydrocarbon and non-hydrocarbon gases found in porous rock formations below the earth's surface, usually with crude oil. The main component of natural gas is methane, together with small amounts of heavier hydrocarbons and some non-hydrocarbon gases, such as carbon dioxide (CO_2), helium, hydrogen sulfide (H_2S) and nitrogen (N_2). It can further be classified as:

Associated gas: the gas occurs in the free state in contact with a large crude oil reservoir. Associated gas usually contains less methane than nonassociated gas, but is richer in higher molecular weight hydrocarbons.

Nonassociated gas: the gas remains in the free state, associated with minimal amounts of crude oil. It is richer in methane and leaner in other higher molecular weight hydrocarbons than associated gas.

Dissolved gas: the pressure in the oil reservoir is high, so the gas remains in solution. It is released as a result of drilling. Natural gas can be in other ways classified as: dry or lean gas which contains mostly methane (CH_4); wet gas which contains considerable quantities of higher hydrocarbons; sour gas which contains much H_2S; sweet gas which contains little H_2S; residue gas from which higher paraffins have been extracted, and finally casing head gas which is derived from an oil well by extraction at the surface.

The non-hydrocarbon components of natural gas can be classified as:

- Diluents such as N_2, CO_2 and water vapor
- Contaminants such as H_2S and other sulfur compounds.

The diluents, being non-combustible, reduce the value of the gas as a fuel and thus can be used as fillers to reduce its heat content. The contaminants can cause harmful effects and can also affect transportation and production equipment plants.

Natural Gas Liquids (NGL): Liquid hydrocarbon mixtures extracted from natural gas. They fractionate into ethane (C_2H_6), propane (C_3H_8) and butane (C_4H_{10}). Liquefied petroleum gas (LPG) containing mainly C_3H_8 and C_4H_{10} is the main commercial product made from natural gas and it can be stored and handled as a liquid at ambient temperatures and moderate pressures.

Natural gasoline is a mixture of hydrocarbons extracted from natural gas mostly containing pentanes and some heavier hydrocarbons. Initially, liquefied petroleum gas and natural gasoline were used solely as fuels, but recently liquefied petroleum gas has been used as a source of propane and butane for the petrochemical industry. Natural gasoline is also used as a source of butanes (C_4H_{10}), isobutanes

(iC_4H_{10}), pentanes (C_5H_{12}) and isopentanes (iC_5H_{12}) for reforming, alkylation, synthetic rubber manufacture, *etc*. Table 11-16 shows the general composition of natural gas.

Table 11-16 Composition of natural gas.*

Source: Majoorana, S., K. B., Modern Petroleum Chemistry, Kannatheri Publications, 2003.

Component	Dry gas (mol %)	Sour gas (mol %)	Gas condensate (mol %)
CO_2	–	6.7	0.7
H_2S	–	3.3	–
N_2 and air	0.8	–	–
CH_4	95.8	84.0	74.5
C_2H_6	2.9	3.6	8.3
C_3H_8	0.4	1.0	4.7
iC_4H_{10}	0.1	0.3	0.9
nC_4H_{10}	Trace	0.4	1.9
iC_5H_{12}	–	–	0.8
nC_5H_{12}	–	–	0.6
C_6H_{14}	–	0.7	1.3
C_7H_{16}	–	–	6.3
*Sometimes natural gas contains some helium content 0–0.5 mol %.			

11.15.2. Gas Processing Methods

There are many gas processing technologies as described in Hydrocarbon Processing Magazine and Handbook (see: www.HydrocarbonProcessing.com). The main process types are:

- Reaction type: Monoethanolamine (MEA), diethanolamine (DEA), methyldiethanolamine (MDEA), diglycoamine (DGA)
- Physical solvent type: Flour solvents (propylene carbonate), Selenol

- Physical/chemical type: Sulfinol.
- Carbonate type: Potassium carbonate
- Solution batch type: Lo-Cat, Chemsweet
- Bed batch type: Iron Sponge, Molecular sieve.

Figure 11-14 provides a guide line for selecting gas processes.

Figure 11-14

Guide for selecting gas sweetening processes

(Source: Branan, Carl, R., "Rules of Thumb for Chemical Engineers" Gulf Publishing Co., 1994).

11.15.3. Reaction Gas Processes

The amines (MEA, DEA, MDEA and DGA) are the most popular treating solutions under the following conditions:

- Low acid and gas partial pressures (product of system pressure and concentration of acid gases H_2S and CO_2 in the feed) of roughly 50 psia and below.
- Low acid gas concentrations in the product of roughly 4–8 ppmv (0.25–0.5 grains/100 scf).
- Heavier hydrocarbons present (provide better filtration for DGA).
- For CO_2 removal with no H_2S, a CO_2 partial pressure of 10–15 psi [308].

Among the amines, MEA is preferred for low contactor pressures and stringent acid gas specifications such as H_2S well below 0.25

grains/100 scf, and CO_2 as low as 100 ppmv. MEA is degraded by carbonyl sulfide (COS) and carbon disulfide (CS_2), with the reactions being only partially reversible with a reclaimer. (The reclaimer helps remove degradation products from the solution and also aids in the removal of suspended solids, acids and iron compounds. Reclaimer operation is a semi-continuous batch operation, where it is filled with hot amine solution and if necessary, soda ash is added.)

DEA is preferred when the total system pressure is >500 psi or where pipeline quality gas is not required. The 0.25 grains/100 scf constraint is more difficult to reach with DEA. COS and CS_2 have few detrimental effects on DEA.

The process flow scheme for DEA plants resembles that for the MEA process, however the advantages and disadvantages of DEA compared to MEA are:

The mole/mole loading typically used with DEA (0.35–1.3 mole/mole) is much higher than that used (0.3–0.4) for MEA.

- Because DEA does not form significant amounts of non-regenerable degradation products, a reclaimer is not required.
- DEA is a secondary amine, and so is chemically weaker than MEA, hence less heat is required to strip the amine solution.
- DEA is typically used for refinery and manufactured gas streams that have COS and CS_2 present.

MDEA is preferred for selective H_2S removal and lack of degradation by COS and CS_2.

DGA is preferred for cold climates and high (50–70 wt%) solution strength for economy. The freezing point for 50% DGA solution is –30°F. By comparison, solution strength for MEA is 15–25 wt%, and

for DEA, 25–35 wt%. Because of the high amine degradation rate DGA systems require reclaiming to remove the degradation product.

\MDEA is a tertiary amine which is used to selectively remove H_2S to levels required by pipeline specifications at moderate to high pressure, if increased concentration of CO_2 in the residue gas does not cause problems. Some of the advantages of selective removal of H_2S are:

- Smaller amine regeneration plants are needed.
- Reduced solution flow rates resulting from a reduction in the amount of acid gas removed.
- Enriched H_2S concentrations in the acid gas resulting in reduced problems in sulfur recovery.

11.15.4. Physical Solvent Gas Processes

The physical solvent types of gas processing are considered when acid gases in the feed are above 50–60 psi, heavy hydrocarbon concentration in the feed gas is low, bulk removal of the acid gas is required and high selective removal of H_2S is desired. Usually, physical solvents are economical because regeneration occurs by flashing or stripping which requires little energy. The solvents are regenerated by multistage flashing at low pressures; regeneration at low temperatures with an inert stripping gas; heating and stripping of solution with steam/solvent vapors.

Selenol®: (licensed by Norton Company) uses dimethylether or polyethylene glycol as a solvent. A Selenol® plant can be designed to provide some selectivity for H_2S, *e.g.* the plant can be designed to provide pipeline quality gas (0.25 grains H_2S/100 scf) while slipping 85% of the CO_2 present in the original sour gas.

The Fluor Solvent: propylene carbonate is used primarily for removal of CO_2 from high pressure gas streams. The temperature of the lean

solution in the absorber is usually well below ambient, and some method of refrigerating the solvent is required. Propylene carbonate is not economical below about 12% acid in the feed [284].

Rectisol: This method, developed by Lurgi Co. and Linde A.G, uses methanol as a solvent. Because of the vapor pressure of methanol, the process is normally applied at temperature of –30°F to –100°F. The process has been found suitable for the purification of natural gas for LNG production. It is best suited to situations where there is little ethane and heavier components in the feed.

***Purisol*:** The process has been developed by Lurgi, and licensed by the Ralph Parsons Company. The solvent used in this process is N-methyl-2-pyrolidone (NMP), a high boiling point liquid.

11.15.5. Physical and Chemical Processes

The Sulfinol process developed by Shell Development Company uses both physical and chemical processes for gas treatment. It blends a physical solvent and an amine to obtain the advantages of both. The physical solvent is Sulfolane® (tetrahydrothiophene dioxide) and the amine is usually DIPA (diisopropanol amine). The flow scheme is identical to an amine plant.

11.15.6. Advantages and disadvantages of the Sulfinol® process

Advantages

- The solvent Sulfolane®, like other solvents, has higher capacity for acid gas at higher acid gas partial pressures. At these higher partial pressures, the Sulfinol® process has lower circulation rates (higher solution loading) and better economy than MEA. Sulfinol® is preferred over MEA when the H_2S/CO_2 ratio is >1:1, while at high acid gas partial pressures.
- In solution, the amine DIPA is able to achieve pipeline quality gas (0.25 grains H_2/100 scf).
- COS, CS_2, and mercaptans (RSH) are removed. CO_2 slightly

DIPA degrades, but reclaimation is easy.

- Low corrosion/carbon steel.
- Low foaming.
- Low vapor losses.

Disadvantages

- Sulfinol absorbs heavy hydrocarbons, which may provide problems if the acid gas is then fed to a sulfur recovery unit.
- The process uses expensive chemicals and the user must pay a process royalty, but losses are low and the licensee receives many engineering services.

11.15.7. Carbonate Process

The basic process was developed by the U.S Bureau of Mines and uses an aqueous solution of potassium carbonate (K_2CO_3). The contactor and still both operate at temperatures in the range of 230–240°F. The process is not suitable for gas streams containing only H_2S. Natural gas streams have been economically treated with potassium carbonate. Medium to high acid gas concentrations, combined with the high pressures of natural gas transmission lines yield the high partial pressures of acid gases required for the process.

The overall reaction for CO_2 and H_2S with potassium carbonate can be represented by:

(11-27)

(11-28)

For stringent specifications of outlet H_2S and CO_2, special designs or a two-stage process may be required. Maddox [309] states that a two-stage process would typically be used for acid gas feed concentrations of 20–40%. Maddox gives the following rules of thumb for process flow scheme selection:

- Single-stage process will remove CO_2 down to 1.5% in the treated gas.
- A split stream cooling modification will produce 0.8% CO_2.
- Two-stage process or two-stage process with cooler will go below 0.8% CO_2.

He further shows how the major process concerns of corrosion, erosion and column instability must be met in the design and operation of a hot carbonate process, since these items will affect the capital and operating/maintenance costs.

Various processes such as the Benfield, Caracara and Giammarco processes attempt to improve on the basic potassium carbonate process by using activators to increase the rate of CO_2 absorption.

11.15.8. Solution Batch Process

The Lo-Cat® process can be used to sweeten or convert H_2S to sulfur. It catalytically converts H_2S in the gas stream to solid sulfur and will not remove CO_2, COS, CS_2, or mercaptans. Iron is held in dilute solution (high circulation rates) by the common chelating agent EDTA (ethylene diamine tetra acetic acid). The iron oxidizes the H_2S to sulfur. The solution is circulated batch-wise to an oxidizer for regeneration.

Chemsweet, from C.E. Natco, is another H_2S only process. It uses an aqueous dispersion of zinc oxide and zinc acetate to oxidize H_2S and

form zinc sulfide. The process can handle H_2S up to 100 grains/100 scf at pressures from 75–1400 psig. Mercaptan concentrations above 10% of the H_2S concentration can cause problems by forming zinc mercaptides [Zn(SH)RH]. The resulting zinc mercaptide will form sludge and possibly cause foaming problems.

Process Selection

The treatment of gas can affect the design of the entire gas processing facility, including the methods chosen for sulfur recovery and acid gas removal, liquids recovery, liquids fractionation and liquid product treating. Designers should consider the following factors when selecting a gas treatment process:

- Specification of the acid gas.
- Hydrocarbon composition of the gas.
- Type and concentration of impurities in the sour gas.
- Temperature and pressure at which the sour gas is available, and at which the sweet gas must be delivered.
- Volume of gas to be processed.
- Specification of the residue gas.
- Selectivity required for acid gas removal.
- Air pollution regulations regarding sulfur removal and/or tail gas cleanup (TGCU).
- Capital and operating costs.
- Liquid product specifications.
- Royalty costs for the process.

Process selection and economics depend on knowing all components present in the gas. Impurities such as COS, CS_2 and mercaptans can present significant impact on the design of both the gas treating and downstream processing facilities. Further, if the gas processing equipment is to be used together with the recovery of liquids, then the requirements for H_2S, CO_2, and mercaptans removal may be affected.

With sulfur recovery, the composition of the acid gas stream feeding the sulfur plant must be considered. CO_2 concentration >80% in the acid gas may require selectively treating the gas in order to raise the H_2S concentration to the sulfur recovery plant.

High water concentration and hydrocarbons in the feed gas can cause design and operational problems for the sulfur recovery plant, and therefore the effect of these components must be considered when selecting the gas treatment process. Further decisions are simplified by considering gas composition and operating conditions. For example, high partial pressures (50 psi) of acid gases increase the likelihood of using the physical solvent method. Alternatively, the presence of significant quantities of heavy hydrocarbons in the feed prevents the use of physical solvents. Generally, low partial pressures of acid gases and low outlet specifications would require the use of amines for adequate treatment. Table 11-17 provides a summary for a number of processes.

Table 11-17 Process Capabilities for Gas Treating

Source: Gas Processors Suppliers Association, Engineering Data Book, Vols. 1 and 2, 12th Ed., (2004).

	Normally capable of meeting ¼ grains[††] H_2S	**Removes mercaptans and COS sulfur**	**Selective H_2S removal**	**Solution degraded (by)**
Monoethanolamine	Yes	Partial	No	Yes (COS, CO_2, CS_2)
Diethanolamine	Yes	Partial	No	Some (COS, CO_2, CS_2)
Diglycolamine®	Yes	Partial	No	Yes (COS, CO_2, CS_2)
Methyldiethanolamine	Yes	Slight	Yes[‡]	No

Sulfinol®	Yes	Yes	Yes‡	Some (CO_2, CS_2)
Selenol	Yes	Slight	Yes‡	No
Hot Pot-Benfield	Yes*	No+	No	No
Fluor Solvent	No**	No	No	No
Iron Sponge	Yes	Partial	Yes	–
Mol Sieve	Yes	Yes	Yes‡	–
Stretford	Yes	No	Yes	CO_2 at high concentrations
Lo-Cat	Yes	No	Yes	CO_2 at high concentrations
Chemsweet	Yes	Partial for COS	Yes	No
*Hi–pure version.				
+Hydrolyzes COS only.				
‡Some selectivity is exhibited by these processes.				
**Can be met with special design features.				
‡‡¼ grain H_2S /100 scf				**ppm H_2S**

Chemical Reactions

The overall chemical equilibrium reactions applicable for H_2S and CO_2 reacting with primary and secondary amines are:

For hydrogen sulfide H_2S removal:

(11-29)

(11-30)

(11-31)

(11-32)

(11-33)

(11-34)

For carbon dioxide (CO_2) removal:

(11-35)

(11-36)

(11-37)

(11-38)

(11-39)

(11-40)

The above reactions proceed to the right at high pressures and/or low temperatures and to the left at high temperatures and/or low pressures. Secondary amines undergo similar reactions to primary amines. Tertiary amines can only react with CO_2 via the acid/base reaction in Equation 11-29. The CO_2 reaction is slowed by the time required to dissolve CO_2 and its conversion to bicarbonate. This may be the reason that tertiary amines exhibit a greater selectivity for H_2S in the presence of CO_2.

Process Description

The process flow diagram for an alkanolamine treating plant is shown in Figure 11-15.

Figure 11-15

Typical gas sweetening by chemical reaction.

Here, the sour gas enters through an inlet separator for the removal of liquids and/or solids. From the separator, the gas stream enters the

bottom of the contactor where it contacts the amine solution flowing down from the top of the column. The acid gas components in the gas react with the amine to form a regenerable salt. As the gas continues to pass up the contactor, more acid gases chemically react with the amine. The sweetened gas leaves the top of the contactor and passes through an outlet separator to trap any solution which may be carried over. Where MEA is the sweetening agent, or the contactor is operating at unusually high temperatures, a water wash may be used to attempt to recover some of the vaporized amine from the gas leaving the contactor. If a water wash is used, it consists of three to four trays at the top of the contactor being used as the absorption section, with make-up water to the unit being used as the wash liquid. The sweet gas exiting the contactor is saturated with water, so dehydration is required prior to the sale of gas. The amine-rich solution leaving the contactor flows through a flash tank to remove absorbed hydrocarbons. From there, the solution passes through the rich/lean exchanger where heat is absorbed from the lean solution. The heated, rich amine then passes to the top section of the stripper. As the solution flows down the column to the reboiler, it is stripped of H_2S and CO_2. The amine solution then passes through the rich/lean exchanger and a solution cooler, to reduce the lean solution temperature to approximately 10°F above that the inlet gas in order to stay above the hydrocarbon dew point. The process is repeated as the lean solution is returned to the contactor.

Finally, acid gas stripped from the amine exits from the top of the stripper, where it passes through a condenser to cool the stream and to recover water. The recovered water is usually returned to the stripper as reflux. The acid gas passes through an outlet separator and, depending on composition and quantity, is vented, sent to sulfur recovery plants, or compressed for sale.

11.15.9. Process Description of Gas Processing Using UniSim® Simulation

Software

A Case Study

A natural gas stream containing N_2, CO_2 and C_1–nC_4 is processed in a refrigeration system to remove the heavier hydrocarbons. The lean, dry gas produced will meet a pipeline hydrocarbon dew point specification. The liquids removed from the rich gas are processed in a depropanizer column, yielding a liquid product with specified propane content.

UniSim® simulator (gas-processing USC) is used to model a natural gas processing facility that uses propane refrigeration to condense liquids from the feed, and a distillation tower to process the liquids. Figure 11-16 show a snapshot of the flowsheet, and Figures 11-17 shows the phase envelope of the feed having the bubble and the dew points. Figure 11-18 shows the hydrate line intersecting the phase envelope. The critical temperature and pressure of the combined feed are 12.28°C and 9976 kPa respectively. The combined feed stream enters an inlet separator, which removes the free liquids. Overhead gas from the SEPARATOR is fed to the gas/gas exchanger, where it is pre-cooled by already refrigerated gas. The cooled gas is then fed to the chiller, where further cooling is accomplished through exchange with evaporating propane (e.g. C3 Duty stream). In the chiller, which is modeled as a COOLER, enough heavier hydrocarbons condense such that the eventual sales gas meets a pipeline dew point specification. The cold stream is then separated in a low temperature separator (LTS). The dry cold gas is fed to the gas/gas exchanger and then to sales, while the condensed liquids are mixed with free liquids from the inlet separator. These liquids are processed in a depropanizer column to produce a low propane content bottoms product. Figure 11-19 and Figure 11-20 show the phase envelope of the sales gas and with the hydrate line outside the boundary of the phase envelope having critical temperature and pressure as –33.93°C and 8070 kPa. Results of simulation of natural gas processing (gas-processing. HSC)

are shown in Appendix L.

Figure 11-16

Process flow diagram of a gas processing unit.

Figure 11-17

Phase envelope of natural gas feed.

Figure 11-18

Phase envelope of natural gas feed.

Figure 11-19

Phase envelope of Sales gas.

Figure 11-20

Phase envelope of Sales gas.

Simplified Design Calculations for MEA, DEA and DGA®

The following equations are used for making rough estimates of the principal parameters for conventional MEA, DEA and DGA® amine treating facilities, when both H_2S and CO_2 are both present in the gas. The procedure involves calculating the amine circulation rate and using it as the principal variable in estimating other parameters [290].

For MEA:

(11-41)

In S.I. units:

(11-41a)

(0.33 mol acid gas pick-up per mole MEA assumed)

For DEA (conventional):

(11-42)

(11-42)

In S.I. units:

(11-42a)

(0.5 mol acid gas pick-up per mole DEA assumed)

For DEA (high loading):

(11-43)

In S.I. units:

(11-43a)

(0.7 mol acid gas pick-up per mole DEA assumed)

For DGA®:

(11-44)

In S.I. units:

(11-44a)

(0.39 mol acid gas pick-up per mole DGA® assumed)

(DGA® concentrations are normally 50-60% by weight)

where

Q = Sour gas to be processed, Mescd, MSm^3/day
x = Amine concentration in liquid solution, wt%, mass %
y = Acid gas concentration in sour gas, mole %

After the amine circulation has been estimated, heat and heat exchange requirements can be calculated from the information in Table 11-18; the pump power requirements can be estimated from Table 11-19.

Table 11-18 Estimated Heat Exchange Requirements

Source: Gas Processor Suppliers Association, Engineering Data Book, Vols. 1 and 2, 12th Ed., (2004).

	Duty		**Area**	
	Btu/hr	**kW**	**Sq. ft.**	**m^2**
Reboiler (Direc fired)	72,000 • GPM	93 • m^3/h	11.30 • GPM	4.63 • m^3/h
Rich-Lean Amine HEX	45,000 • GPM	85 • m^3/h	11.25 • GPM	4.60 • m^3/h
Amine cooler (air cooled)	15,000 • GPM	19.3 • m^3/h	10.20 • GPM	4.18 • m^3/h
Reflux condenser	30,000 • GPM	38.6 • m^3/h	5.20 • GPM	2.13 • m^3/h

Table 11-19 Estimated Power Requirements

Source: Gas Processor Suppliers Association, Engineering Data Book, Vols. 1 and 2, 12th Ed., (2004).

Main Amine Solution Pumps	GPM • PSIG • 0.00065 = hp	m^3/h • kPa (ga) • 0.00031 = kW
Amine Booster Pumps	GPM • 0.06 = hp	m^3/h • 0.20 = kW
Reflux Pumps	GPM • 0.06 = hp	m^3/h • 0.20 = kW
Aerial Cooler	GPM • 0.36 = hp	m^3/h • 1.20 = kW

Equations (11-41) and (11-41a) provide conservative (high) estimates of required circulation rates. They should not be used if the combined H_2S and CO_2 concentration in the gas is above 5 mole %. They also are limited to a maximum amine concentration of about 30% by weight.

The diameter of an amine plant contactor is:

(11-45)

In S.I. units:

(11-45a)

where

D_c = Contactor diameter in inches (before rounding up to the nearest 6 inches), or contacting diameter in mm (before rounding up to nearest 100 mm).

Q = MMscfd gas to contactor, MSm^3/day gas to contactor.

P = Contactor pressure, psia, kPa(abs).

The diameter of the regenerator is:

(11-46)

In S.I. units:

(11-46a)

where

D_r = Regenerator bottom diameter in inches, mm.

The diameter of the section of the still above the feed point can be estimated as 0.67 times the bottom diameter.

Example 11-2

40.0 MMscfd of gas available at 900 psig, and containing 0.5% H_2S and 2.5% CO_2 is to be sweetened using 25% by weight DEA solution. If a conventional DEA system is to be used, what amine circulation rate is required, and what will be the principal parameters for the DEA treatment system?

Solution

Using Equation 11-42, the required solution circulation is:

Heat exchange requirements are shown in Table 11-18

Reboiler

$H = 72{,}000 \times 216 = 15.6 \times 10^6$ Btu/hr

$A = 11.30 \times 216 = 2441$ ft^2

Rich-Lean amine exchanger

$H = 45{,}000 \times 216 = 9.72 \times 10^6$ Btu/hr

$A = 11.25 \times 216 = 2430$ ft^2

Amine cooler

$H = 15{,}000 \times 216 = 3.24 \times 10^6$ Btu/hr

$A = 10.2 \times 216 = 2203\ ft^2$

Reflux condenser

$H = 30{,}000 \times 216 = 6.48 \times 10^6$ Btu/hr

$A = 5.20 \times 216 = 1123\ ft^2$

Power requirements from Table 11-19

Main amine pumps, hp = 216 × 900 × 0.00065 = 126.4

Amine booster pumps, hp = 216 × 0.06 = 13

Reflux pumps, hp = 216 × 0.06 = 13

Aerial cooler, hp = 216 × 0.36 = 78

Contactor diameter from Equation 11-45

Regenerator diameter below feed point:

Regenerator diameter above feed point:

$D_{ra} = 0.67 \times 48 = 32.2$ inches or 36 inches (top) rounded up to nearest 6 inches.

11.16. Gas Dryer (Dehydration) Design

Natural gas and associated condensate are often produced from a reservoir saturated with water. Additionally, the gas and condensate often contain CO_2 and H_2S which require removal. Also, many sweetening agents employ aqueous solutions for treating the gas. Dehydration is the process used to remove water from natural gas and natural gas liquids (NGLs) and it is necessary to:

- Prevent the formation of hydrates and condensation of free water
- Prevent corrosion
- Meet a water content specification.

Dehydrating the natural gas after the sweetening process involves:

- Dehydrating by refrigeration
- Dehydration with $CaCl_2$
- Dehydration by distillation
- Dehydration by membrane permeation
- Dehydration by gas stripping
- Absorption by liquid desiccants
- Adsorption by solid desiccants.

Adsorption is a process that involves the transfer of a material from one phase to a surface where it is bound by intermolecular forces, and involves the transfer from a gas or liquid to a solid surface. It could also involve transfer from a gas to a liquid surface. The adsorbate is the material being concentrated on the surface and the material that it

accumulates is known as the adsorbent.

The oil and chemical industries use adsorption processes in the cleanup and purification of wastewater streams, dehydration of gases and in gas purification to remove sulfur dioxide from a stack gas. In addition, adsorption is also employed to fractionate fluids that are difficult to separate by other methods. Porous desiccants (adsorbents) with a large internal surface area are used for industrial applications, as the amount of adsorbate collected per unit area is very small.

The most common commercially use desiccants are alumina, silica gel, and molecular sieves. The last of which posseses the highest water capacity, will give the lowest water dew points, and can be applied to sweeten dry gases and liquids.

Figure 11-21 shows how desiccants can be used in dehydrators containing two or more towers and Table 11-20 presents the most important properties of commercial solid desiccants. Figure 11-21a shows a typical geometry of solid desiccant dryer. One of the towers is in operation, using steam to adsorb water from gas, while the other tower is regenerated and cooled. The regenerated gas is heated to a range of 450°F to 600°F, depending on the type of solid desiccant and the nature of service [292].

Figure 11-21

Solid desiccant dehydrator twin tower system.

Table 11-20 Typical Desiccant Properties

Source: Engineering Data Book: Gas Processor Suppliers Association, 12th Ed., Tulsa, Oklahoma, 2004

Desiccant	Shape	Bulk density, lb/ft^3	Particle size	Heat capacity, Btu/(lb. °F)	Approx. minimum moisture content of effluent gas
Alumina Alcoa F200	Beads	48	7×14 Tyler mesh 1/8” 316” / ¼”	0.24	−90°F dew point
Activated Alumina UOP A-201	Beads	46	3-6 mesh or 5-8 mesh	0.22	5–10 ppmv
Mol Sieve Grace – Davison 4A	Beads	42–45	4-8 mesh or 8-12 mesh	0.23	0.1 ppmv (−150°F)
Molecular Sieve UOP 4A-DG	Extrudate	40–44	1/8” or 1/16”	0.24	0.1 ppmv
Mole Sieve Zeochem 4A	Beads	45–46	4-8 mesh or 8-12 mesh	0.24	0.1 ppmv
Silica Gel Sorbead®-R	Beads	49	5×8 mesh	0.25	−60°F dew point
Silica Gel Sorbead® -H	Beads	45	5×8 mesh	0.25	−60°F dew point
Silica Gel Sorbead($)-WS	Beads	45	5×8 mesh	0.25	−60°F dew point

Figure 11-21a Geometry of a typical solid desiccant dryer.

The use of solid desiccant is limited to applications with very low water dewpoint requirements, where simultaneous control of water and hydrocarbon dewpoints are required, and in very sour gases. In cryogenic plants, solid-desiccant dehydration is much preferred over methanol injection, as it prevents hydrate and ice formation.

The basic design procedures for dehydrating a saturated natural gas at a specified dew point are:

1. Determine the process conditions and the dryer process flow diagrams.
2. Select the drying cycle and calculate the water load.
3. Select the desiccant type and compute the capacity and volume required.
4. Size the dryer and check for pressure drop.
5. Calculate the desiccant reactivation heating and cooling requirements.

Here, design procedures are presented for sizing a solid desiccant dryer for removing moisture from gas streams. The process involves the following calculations.

- The water pickup
- Desiccant volume
- Desiccant bed
- Gas velocity through the bed
- Vessel weight
- Dryer regeneration requirements.

The design is based upon the following assumptions [292]:

1. The temperature difference between the heater outlet temperature and the peak vessel outlet temperature is 50°F.
2. The average bed temperature is based on 75% of the bed at heater outlet temperature and 25% at the peak vessel outlet temperature Equation 11-56.
3. Specific heats: steel, 0.12 Btu/lb°F, and desiccant 0.25 Btu/lb°F.
4. Heat of water desorption 1400 Btu/lb H_2O adsorbed.
5. Flat heads used on vessel ends; steel density is 480 lb/ft^3.
6. Total vessel weight increased by 10 per cent for supports.
7. Heat losses to the dryer heating period calculated at 5%.

11.16.1. The Equations

The equations used for calculating these requirements are:

The total water adsorbed, lb:

(11-47)

Desiccant volume:

(11-48)

The calculated area based on desiccant volume:

(11-49)

The bed diameter:

(11-50)

Actual desiccant bed area:

(11-51)

The actual gas flow rate at flowing conditions, :

(11-52)

Superficial gas velocity, ft/min:

(11-53)

Superficial gas velocities for adequate contact time should range between 30 and 60 ft/min.

Dryer shell thickness, inch:

(11-54)

Vessel weight:

(11-55)

Add an extra 2 ft (h = 2) of straight side (i.e. for 6 in. of 3/8 in. inert balls above and below the desiccant bed and 1 ft for an inlet-gas distributor).

Temperature of desiccant bed, °F:

(11-56)

The total heat required to regenerate the dryer (ideal), Btu:

(11-57)

where:

(11-58)

Loss factor for non-steady state heating, F:

(11-59)

Heat added to regeneration gas to regenerate the desiccant bed, Btu:

(11-60)

The total regeneration gas requirement, lb:

(11-61)

Standard cubic feet of regeneration gas:

(11-62)

11.16.2. Pressure Drop

The pressure drop is estimated from the desiccant manufacturer's data and relevant correlations. In addition, can be calculated by Wunder's [310] graphical correlation of gas superficial velocity against in ft water per ft bed. Wunder used a factor of 1.6 for a fouled bed. through a new desiccant bed is initially low. After a short time in service, rises because of bed settling and then slowly increases over the active life of the desiccant, because of more settling and some attrition. Figure 11-22 shows for an 8-inch silica gel desiccant. These data are based on using air. For other gases, the given values should be multiplied by . The pressure drop for a clean bed is given by:

(11-63)

(11-63)

Figure 11-22

Pressure drop for an 8 mesh silica gel desiccant

(source: Wunder, J. W., "How to Design a Natural Gas Drier", Oil & Gas Journ., Aug. 6, pp. 137-148, (1962)).

11.16.3. Fouled Bed

(11-64)

Desiccant attrition due to high gas velocities is controlled by proper bed design. The superficial gas velocity is used to estimate Alcoa's momentum down flow velocities [310]. However, desiccant attrition should not be a problem when the momentum number is equal to or less than 30,000. Although this number is based on granular alumina, it can be used for silica gel, which has crushing characteristics similar to alumina. The momentum is expressed as:

(11-65)

where

V_{GAS} = superficial gas velocity, ft./min

MW = molecular weight of gas

P = system pressure in psig.

Desiccant Reactivation

The desiccant bed is reactivated at as high a temperature as permissible to regain optimum capacity and minimum dew points. In natural gas drying service, the normal reactivation temperature for silica gel is between 350°F and 400°F. Because the feed gas contains high levels of pentane-plus materials, Wunder [310] suggested that reactivation can be carried out at 450°F to desorb the heavy hydrocarbons.

At a temperature above 450°F, the attrition rate for silica gel is high. However, occasionally at 600°F, reactivation will restore some desiccant's initial activity thus extending the life of the silica gel. Therefore, the reactivation procedure is designed to periodically heat the desiccant to 600°F [310].

Example 11-3

Size a dryer that uses silica gel using the data in Table 11-21.

Table 11-21 Dryer Parameters

Flow = 65 MM std ft^3/day	Desiccant H_2O cap = 8%
P = 700 psig	Desiccant density = 50 lb/ft^3
T = 80°F	Desiccant bed length = 15 ft
Z = 0.9228	Heater outlet = 500°F
H_2O Cont = 47 lb/MM std ft^3	MW = 18
S = 15,500 lb/in^2	H_1 = 490 But/lb
E = 1.0	H_2 = 225 Btu/lb
Cycle = 12h	h = 2.0 ft.

Solution

Computer program PROG 113 has been developed to calculate the total regeneration flow, the heat requirement for dryer regeneration, the momentum and superficial gas velocity. Table 11-22 illustrates the input data and the computer output. The total regeneration flow is 44649.0 lb and the momentum is 30,584. The heat requirement for dryer regeneration is 5.6×10^6 Btu. The calculated superficial gas velocity is 34.9 ft/min.

Table 11-22 Input Data and Computer Output for Gas Dryer Design

Data name: Data113. Dat		
65.0	700.0	80.0
0.9228	12.0	47.0
15500.0	1.0	8.0
50.0	15.0	500.0
18.0	490.0	225.0
Gas Dryer Design		
Gas flow rate, 10^6 std. ft^3/day:	65.000	
Dryer operating pressure, psig.:	700.000	
Gas flowing temperature, °F:	80.0	
Gas compressibility factor:	0.9228	
Drying cycle, hr.:	12.0	
Water content of gas, lb/10^6 std.ft^3:	47.	
Allowable stress, psi:	15500.0	
Joint efficiency:	1.0	
Useful design capacity of desiccant, (%):	8.0	
Desiccant density, 1b/ft^3:	50.0	
Heater outlet temperature, °F:	500.0	
Molecular weight of regeneration gas:	18.0	
Enthalpy of gas before regeneration heater, Btu/1b:	490.0	

Enthalpy of gas before regeneration heater, Btu/1b:	[illegible]
Enthalpy of gas after regeneration heater, Btu/1b:	225.0
Total water adsorbed, 1b/h.:	1527.500
Desiccant volume, ft^3.:	381.8750
Calculated crosssectional area, ft^2.:	25.4583
Bed diameter, ft.:	5.6934
Bed radius, ft.:	34.1603
Bed length, ft.:	15.00
Actual gas flow rate, ft^3/min.:	889.697
Superficial gas velocity, ft./min.:	34.947
Dryer shell thickness, in.:	1.5857
Vessel weight, 1b.:	24767.
Desiccant weight, 1b.:	19094.
Temperature of desiccant bed, °F:	487.500
Total heat required to regenerate dryer, Btu:	0.555954E + 07
Loss factor:	2.1282
Heat added to regeneration gas, Btu:	0.118320E + 08
Total regeneration gas requirement, lb:	0.446490E + 05
Standard cubic feet of regeneration gas, std. ft^3:	0.942590E + 06
Momentum:	0.305838E + 05

From Figure 11-22:

The superficial gas velocity = 35 ft./min.

Desiccant bed length = 15.0 ft.

System pressure = 700 psig.

The clean bed using Equation 11-63 is:

Part 2 Hydrocarbon Absorption and Stripping

(With Contributions by Dr. P. A. Bryant)

Many operations in petrochemical plants require the absorption of components from gas streams into “lean” oils or solvents. The resultant “rich” oil is then stripped or denuded of the absorbed materials. The greatest use of this operation utilizes hydrocarbon materials, but the basic principles are applicable to other systems provided adequate equilibrium data are available.

Several methods [17, 18, 29, 40, 62, 67, 223] for handling this design have been offered, and each has introduced a concept to improve some feature. An approximation method combining Kremser-Brown [40, 67] and a more complete method of Edmister [18] will be summarized. Figure 11-23 summarizes the system and terminology.

11.17. Kremser-Brown-Sherwood Method - No Heat of Absorption

This method [18] gives reasonably good results for systems involving relatively lean gas and small quantities absorbant. For rich gases the error can be considerable (more than 50% for some components). It has given generally good results on natural gas and related systems.

11.17.1. Absorption: Determine Component Absorption in Fixed Tray Tower (Adapted in part from Ref. 18)

1. Calculate the total mols. of gas inlet to the absorber, identifying the quantities of individual components.
2. Assuming the tower pressure as set and that an average of top and bottom temperatures can be selected (these may become variables for study), read equilibrium K_i values from charts for each component in gas.
3. Assume or fix a lean oil rate.

4. Calculate

(11-66)

Assume this is a constant value for tower design.

5. Calculate absorption factor, $A_i = L_o/(V_{N+1})\ (K_i)$, using values of (2) and (4) above for each component.

6. Calculate the fraction absorbed for each component, assuming a fixed overall tray efficiency for an assumed number of actual trays (or an existing column with trays).

a. Theoretical number of trays, N = (tray efficiency, E_o,) (number of actual trays)

b. Fraction absorbed

(11-67)

where Y_o^* is often considered zero or very small. Solve using A_i values.

7. Mols each component absorbed/hr.

8. Mols each component absorbed/(mol inlet lean oil)

9. Mols of each component in gas out of top of absorber:

= (mols component in inlet gas) − (mols component absorbed)

10. Mols of component in gas out of top of absorber/mol of inlet rich gas:

Solve for Y_1.

11. Correct values from first calculation, Steps 1 through 10, using the ΣX_{iR} values of Step 8, as follows:

12. Calculate A_i:

(11-68)

13. Calculate absorption efficiency, E_{ai}. Using new A_i value:

read Figure 11-24

14. Calculate mols absorbed/hr:

= (E_{ai}) (mols component in inlet rich gas)

15. Mols of each component in gas out of top of absorber/hr = (mols component in) − (mols component absorbed)

16. Mols of component in outlet gas/mol inlet rich gas. Solve for Y_{1i}

If the X_1 in equilibrium with Y_1 is desired:

(11-69)

17. Improved values can be obtained by recalculation from Step 11 if there is too great a difference between the "Σ mols absorbed" from trial no. 1 and trial no. 2.

First trial

Component	Inlet $Y_{(N+1)i}$	Mols/hr In	K_i	
•	•	•	•	•
•	•	•	•	•
•	•	•	•	•
	1.00	Σ •		
Component	**Fraction Absorbed E_{ai}**	**Mols Absorbed**	**X_{iR}**	**Mol/hr Off Gas**
•	•	•	•	•
•	•	•	•	•
•	•	•	•	•
		Σ •	Σ •	Σ •

For second trial, see Step 11.

18. A graphical stepwise is procedure offered by Sherwood [62] and also summarized by Reference 18. Y and X are plotted and handled stepwise as in distillation. The equilibrium line equation is, for any single component:

(11-70)

For a complete denuded inlet solvent at the top ΣX = 0, using K at top column conditions. The slope of the operating line = L_o/V_{N+1} = mols lean oil entering/mols wet gas entering.

11.17.2. Absorption: Determine the Number of Trays for Specified Product Absorption

1. For fixed tower temperature, pressure, gas feed rate, specified

or assumed operating (L_o/V_{N+1}) times minimum value, specified component recovery from inlet gas.

2. Calculate:

a. Mols component in inlet gas/hr

b. Mols in outlet gas

c. Mols component absorbed = inlet − outlet

3. Calculate: E_{ai} for specified component (specified in 1.)

= specified fraction recovery

4. Minimum (L/V) for specified component:

Assuming equilibrium at bottom, $\Sigma Y = 1$. Ignoring ΣX gives slightly conservative value,

5. Operating $(L_o/V_{N+1})_o$

= (specified unit) $(L_o/V_{N+1})_{min}$

6. Operating

(11-71)

7. Theoretical number of plates at operating (L_o/V_{N+1}): solve for N.

(11-72)

8. Actual number of trays at operating (L_o/V_{N+1}): $N_o = N/E_o$

E_o values may be calculated from Figure 10-54 or assumed at 20 to 50% as an estimating value for hydrocarbon oil and vapors, pressures atmospheric to 800 psig, and temperatures of 40°F to 130°F (see Table 11-23).

Table 11-23 Absorption-Stripping Approximation Tray Efficiencies**

Type	Pressure range, psig	Temp., °F	Range efficiency %
Absorption			
Hydrocarbon Oils & Vapors	0–800	30–130	35–50
Propane-key	100–2,100	–	30–37*–38
Butane-key	100–2,100	–	28–33*–36
Stripping			
Hydrocarbon Oils with Steam	0–130	300–550	50–80
Unsaturates in Oil with closed reboiler	0–50	–	25–35
*Average value			
**Hull, R. J. and K. Raymond, *Oil and Gas Journal*, Nov. 9, 1953 – March 15, 1954 [32], used by permission.			

9. Lean oil rate:

(11-73)

10. For other components: E_{ai} is estimated by

(11-74)

with a limiting value of unity.

11.17.3. Stripping: Determine the Number of Theoretical Trays and Stripping Steam or Gas Rate for a Component Recovery [18]

The rich gas from the absorption operation is usually stripped of the desirable components and recycled back to the absorber (Figure 11-23). The stripping medium may be steam, or a dry or inert gas (methane, nitrogen, carbon oxides, hydrogen, etc.). This depends upon the process application of the various components.

1. The rich oil flow rate and absorbed component compositions (this is the only composition of concern, not the oil composition, unless reaction or change takes place under the system conditions) are known. From the temperature levels of the available condensing fluids (water, refrigerant, etc.), determine a column operating pressure which will allow proper condensation of the desirable components at the selected temperature, allowing for proper temperature difference for efficient heat transfer. The condensing pressure (and column operating pressure) may be dictated by the available steam pressure used in stripping, or the pressure on the inert stripping gas.
2. From the K charts, determine K_i values for each component at the column temperature and pressure.
3. From a fixed percentage of recovery for key component (= E_{si} for key component), mols component stripped/hr = G_{mi} = (L_{m+1})

(X_{m+1}) (E_{si})

4. Estimate stripping efficiency for components other than the key by:

(11-75)

Assume known			ΣX_i =	
V_o	**L_o**	**V_o/L_o**	**$(LX_i)_{m+1}/L_o$**	**1+Σ**
–	Same	–	–	–
–	for	–	–	–
–	all	–	–	–
–	trials	–	–	–
ΣY_i =				
ΣG_{mi}(Step 3)/V_o	**1 + ΣY_i**	**K_{key} (1 + ΣY_i)**		
–	–	–	–	
–	–	–	–	
–	–	–	–	

Note that no recovery can be greater than 1.0, therefore any value so calculated is recorded as 1.0, indicating that the component is completely stripped from the rich oil. Calculate mols stripped per hour for each component as in Step 2.

5. The minimum stripping medium (steam or gas) to lean oil ratio is estimated by a trial and error procedure based on key component: By assuming several values of V_o, plot V_o/L_o versus $E_{s,\ Key}$ $(1 + \Sigma X_i)/K_{Key}$ $(1 + \Sigma Y_i)$. The point where they are equal gives the minimum value for V_o/L_o. This calculation can be

thought of as assuming equilibrium at the gas outlet end and being slightly conservative by including the $(1 + \Sigma X_i)$ term. Operation at this point requires infinite number of plates; therefore, values larger than the minimum should be used. For economical as well as reasonable operation, several values of $(V_o/L_o)_{oper}$ should be tried and corresponding plates evaluated.

V_o (operating) = (assumed $(V_o/L_o)_{oper}$) (L_o inlet), mols/hr

6. Calculate S_i for the key component, using the value of $(1 + \Sigma X_i)$ calculated in Step 5. Calculate

(11-76)

Sometimes the last term on the right can be neglected.

7. Calculate number of theoretical trays, M.

(11-77)

(11-78)

8. Actual number of trays at operating reflux:

(11-79)

9. Calculate for each component corrected amount stripped:

For each component:

(11-80)

10. From Figure 11-24, read $(S^{M+1} - S)/(S^{M+1} - 1) = E_{si}$ for each component at the fixed theoretical required trays and at each individual S_i value.

Figure 11-24

Absorption and stripping factors, E_a or E_s vs. effective values A_e or S_e (efficiency functions).

Used by permission, Gas Processing Suppliers Association, Engineering Data book, vol. 12, 12th ed., Tulsa, Oklahoma (2004).

11. For final detail, recalculate mols stripped per hour from new E_{si} values and the total quantities of each component in the incoming rich oil. If values do not check exactly, adjustments can be made in steam rate and ΣY_i to give exact values. In many cases this accuracy is not justified since the method is subject to some deviation from theoretically correct values.

12. A graphical solution is presented by Edmister [18] and handled like stepwise distillation.

Figure 11-23

Flow diagram of absorption-stripping for hydrocarbon recovery from gaseous mixture.

Used by permission, Edmister, W. C., Petroleum Engr., Sept. (1947) to January (1948).

Equilibrium line: starts at origin of X–Y plot.

For assumed X values, calculate Y corresponding for key component from

At lean oil end of tower: $\Sigma X_i = 0$ and $\Sigma Y_i = 0$.

Slope of equilibrium line is $Y/X = K_i$

At rich oil end of tower:

(11-81)

where X_i, ΣX_i and ΣY_i are known. R = rich end.

Operating line:

At lean end, $Y_i = 0$ (or nearly so in most cases); if not, plot accordingly.

11.17.4. Stripping: Determine Stripping-Medium Rate for a Fixed Recovery [18]

1. The composition and quantity of rich oil, and percent recovery of a specified key component are known, also column pressure and temperature.
2. Using Figure 11-24, assume a value for theoretical plates, read S_e corresponding to specified value of recovery E_{si} for key component, since:

Note that with this procedure, the effect of the number of theoretical plates available can be determined. In an existing column where the number of trays is fixed, the theoretical trays can be obtained by evaluating an efficiency for the system.

3. The value of $S_e = S_i$ for key component obtained in Step 2 is equal to:

Using key component:

Set up table: use K_i for each component to calculate Column 4.

Component	**Mols/hr in Rich Oil**	**K_i at Col. Condition**	
–	–	–	–
–	–	–	–

	Σ		

From values of S_i calculated ($=S_e$), read E_{si} values from Figure 11-24 at the number of theoretical trays assumed in Step 2. Note that the S_e corresponds to the number of trays selected, hence will give a value for performance of the system under these particular conditions.

4. Calculate the mols of each component stripped/hr

5. Calculate V_o, mols/hr of stripping medium required (steam or gas):

for key component.

Multiply by L_o.

This is equal to V_o. Note that V_o also is in the right hand side of the denominator, so fractions must be cleared.

11.18. Absorption: Edmister Method

This method [18] is well suited to handling the details of a complicated problem, yet utilizing the concept of average absorption and stripping factors. It also allows for the presence of solute components in the solvent and the loss of lean oil into the off gas. Reference 18 presents more details than are included here. It is Edmister's original publication of the basic method for absorbers and strippers, and it also generates the treatment to include distillation

towers and presents the same graphical relationships in a slightly modified form.

Absorption: Lean Oil Requirement for Fixed Component Recovery in a Fixed Tower [18]

1. The rich gas is known, the number of theoretical trays is fixed (or assumed and corresponding result obtained), the operating pressure and temperature can be fixed.
2. For key component and its fixed recovery, E_a, read A_e from Figure 11-24 at the fixed theoretical number of trays, N:

3. Assume:
   a. Total mols absorbed
   b. Temperature rise of lean oil (normally 20–40°F)
   c. Lean oil rate, mols/hr, L_o

4. Using Horton and Franklin's [29] distribution relation for amount absorbed (or vapor shrinkage), per tray:

(11-82)

Mols off gas leaving top tray.
$= V_1 = V_{N+1}$ – Mols absorbed (assumed)
Mols gas leaving bottom tray No. N
$= V_N = V_{N+1} (V_1/V_{N+1})^{1/N}$

Liquid leaving top tray No. 1 $= L_1 = L_o + V_2 - V_1$
where
V_2 = vapor leaving tray No. 2 from top, mols/hr

L_o = lean oil entering (assumed completely free of rich gas components), mols/hr L_N = liquid leaving bottom tray, mols/hr V_N = vapor leaving bottom tray, mols/hr

Liquid leaving bottom tray

= $L_4 = L_o$ + Mols absorbed (assumed)

5. Calculate: At top, L_1/V_1

At bottom, L_N/V_N

6. Use Horton-Franklin method to estimate temperatures at tower trays:

(11-83)

where

T_o = lean oil temperature, °F

T_N = bottom tray temperature, °F

T_i = tray, i temperature, °F

T_{N+1} = inlet rich gas temperature, °F

These relations assume constant percent absorption per tray, and temperature change proportioned to the vapor contraction per tray. For estimating use only.

Temperature bottom tray = $T_N = T_{N+1}$ + (assumed rise)

7. Read the K values from equilibrium charts for components in feed at temperatures of (a) top tray and (b) bottom tray.

8. Calculate A_{Ti} and A_{Bi} for each component.

A_{Ti} (for top conditions) = $L_i/(K_iV_i)$ A_{Bi} (for bottom conditions) = $L_N/(K_iV_N)$ where:

A_{Ti} = absorption factor for each component at conditions of top tray A_{Bi} = absorption factor for each component at conditions of

bottom tray

9. Read A_e values corresponding to A_{Ti} and A_{Bi} values from Figure 11-25.

Figure 11-25

Effective absorption and stripping factors used in absorption, stripping and fractionation as functions of effective factors.

Used by permission, Edmister, W. C, Petroleum Engr., Sept. (1947) to Jan. (1948).

10. Read E_{ai} values for fraction absorbed from Figure 11-24 using the A_e values of Step 9 and the fixed or assumed number of theoretical trays.

11. Calculate the mols of each component absorbed by:

(Mol component in inlet rich gas) (E_{ai})

Suggested tabulation:

Component	**Mols Rich**	**K**		**Absorption Factors**			**E_{ai}, Frac.**	**Mols**
	Gas In	**Top**	**Bottom**	**A_T**	**A_B**	**A_e**	**Absorbed**	**Absorbed**
–	–	–	–	–	–	–	–	–
–	–	–	–	–	–	–	–	–
–	–	–	–	–	–	–	–	–

12. If the result does not yield the desired amount of the key component absorbed, then reassume the lean oil quantity, L_o and recalculate. Adjustments may have to be made separately or

simultaneously in the assumed absorption quantity until an acceptable result is obtained. After two or three trials a plot of the key variables will assist in making the proper assumptions.

Absorption: Number of Trays for Fixed Recovery of Key Component

Here we also consider the more general case when the lean oil contains some of the components to be absorbed from the entering gas. The relationships are most conveniently written as follows [18], for a given component:

(11-84)

(11-85)

(11-86)

Rearranging Equation 11-86 gives:

(11-87)

Substituting Equations 11-85 and 11-86 into Equation 11-84 gives

(11-88)

which can be written as

(11-89)

where

v_{1i} = molar gas flow rate of component "i" leaving plate 1 in the absorber

l_{oi} = molar flow of component "i" in entering liquid to the absorber

v_{n+1i} = molar gas flow rate of component "i" in entering gas to the absorber

f_{si} = the fraction of V_{n+li} that is absorbed by the liquid phase

f_{ai} = the fraction of V_{n+1} that is absorbed by the liquid

n = absorber theoretical number of trays, also = N

m = stripper theoretical number of trays, also = M

A material balance on the key component fixes v_1, l_o and v_{n+1} for the key. A_e is estimated after A_T and A_B are calculated using approximate conditions at the top and bottom, with a multiple of the minimum solvent rate, which is estimated by assuming equilibrium at the bottom of the tower. S_e is estimated from S_T and S_B. Note that $S_T = 1/A_T$ and $S_B = 1/A_B$. A trial and error solution for the number of theoretical stages is effected by using Equation 11-89 (or 11-88 and Figure 11-24). Values of v_1 for the non-keys can be calculated by using these relationships directly the calculated value of n; if necessary, the entire procedure can be repeated, using the better estimates of the component flow rates in the leaving streams that were estimated in the first iteration.

Lean Oil

The selection of lean oil for an absorber is an economic study. A light lean oil sustains relatively high lean oil loss, but has the advantage of high mols/gal compared to a heavier lean oil. The availability of a

suitable material has a large influence on the choice. For example, a lean oil three carbon numbers heavier than the lightest component absorbed is close to optimum for some applications. In natural gas plant operations, a lean oil heavier by about 10-14 carbon numbers has been reported [284].

Presaturators

A presaturation to provide lean oil/gas contact prior to feeding the lean oil into the tower is a good way of getting more out of an older tower. Absorber tray efficiencies are low, and therefore a presaturation that achieves equilibrium can provide the equivalent of a theoretical tray. This can easily equal 3-4 actual trays [284].

Example 11-4: Absorption of Hydrocarbons with Lean Oil

The gas stream shown in Table 11-24 is fed to an isothermal absorber operating at 90°F and 75 psia. 90% of the n-butane is to be removed by contact with a lean oil stream consisting of 98.7 mol% nonvolatile oil and the light components shown in column 2 of Table 11-24. Estimate the composition of the product streams and the required number of theoretical stages if an inlet rate of 1.8 times the minimum is used.

Table 11-24 Calculations Summaries for Example

Component	90°F K 75 psia	Feed gas in (mols/hr)	Lean oil in (mol. Fraction)	Initial estimate of net amount absorbed (mols/hr)	Off-Gas		
					Initial estimate (mols/hr)	After first iteration (mols/hr)	After second iteratio (mols/h
Methane	42.5	1,639.2	–	22.6	1,616.6	1,597.5	1,598.4
Ethane	7.3	165.8	–	13.3	152.5	141.2	141.8
Propane	2.25	94.9	–	24.67	70.23	49.84	50.76
i-Butane	0.88	17.8	0.001	11.83	5.97	3.10	3.13

i-Butane	0.88	17.6	0.001	11.83	5.97	5.10	5.15
n-Butane	0.65	33.6	0.002	30.24	3.36	3.36	3.36
i-Pentane	0.28	7.9	0.004	7.9	0	2.08	2.03
n-Pentane	0.225	15.8	0.006	15.8	0	2.51	2.44
Heavy Oil	0	–	0.987	0	0	0	0
		1,975.0	1.000	126.34	1,848.66	1,799.57	1,801.92

Solution

1. Initial estimates of extent of absorption of non-keys.

As a rough approximation, assume the fraction absorbed of a given component is inversely proportional to its K value (Equation 11-74). For example:

The other estimates in column 5 of Table 11-24 are calculated in a similar manner. Note that the C_{5s} are assumed to be completely absorbed for this first iteration.

2. Inlet rate of rich oil.

The maximum mol fraction n-C_4 in the leaving liquid is taken as that in equilibrium with the incoming gases. Thus, for n-C_4,

A material balance of n-C_4 yields:

With the absorption efficiencies assumed above,

Combining the above equations yields the estimate of the minimum lean oil rate:

or $(L_o)_{min}$ = 1,115.3 mols/hr

Thus, L_o = 1.8 (1,115.3) = 2,007.5 mols/hr

3. Effective Absorption Factor for n-C_4.

The total rich oil out is estimated as

L_n = 1,975 + 2,007.5 − 1,848.66 = 2,133.84

The absorption factors are calculated by

A_T = 2,007.5/(.65) (1,848.66) = 1.671

A_B = 2,133.84/(.65) (1,975) = 1.662

The stripping factor, S_e, is taken as $1/A_e$ = 0.60006

4. Calculation of the required number of theoretical stages.

Using Equation 11-88 for n-butane,

which is equivalent to

Solving for n by trial and error yields, n = 5.12

5. Calculation of absorption of non-keys.

Equation 11-89 is used with n = 5.12 to calculate v_1, as for example, for i-butane, $A_T = 1.233$

$A_B = 1.227$

$v_o = 17.8$

$l_o = (0.001)\ (2{,}005.7) = 2.01$

$l_n = 17.8 + 2.01 - 3.097 = 16.71$

The remaining non-keys in the off-gas are calculated in a similar manner and are tabulated in Column 6 of Table 11-24. Note that the calculated values are somewhat different from the assumed values in Column 5.

6. Second Iteration.

Using the previously calculated values, the net amount absorbed is 1,975 − 1,799.59 = 175.41 mols/hr. The minimum rate of lean oil is calculated from

from which $(L_o)_{min}$ = 1,061.2 mols/hr and

$L_o = 1.8\ (1{,}061.2) = 1{,}910.2$ mols/hr

An overall material balance gives L_n = 2,085.6. The effective absorption factor for n-C_4 is $A_e = 1.627$, and $S_e = 0.6145$, n is calculated from

from which n = 5.20 theoretical number of stages.
The non-key components are computed and tabulated in Column 7 of Table 11-24.

7. Third Iteration.

A third iteration gave $(L_o)_{min}$ = 1,063.73, L_o = 1,914.72, L_n = 2,087.8, and V_1 = 1,801.92, with no change in the calculated off-gas component flows.

The stripping calculations are handled in a manner similar to the steps above, and using the figures indicated.

Intercooling for Absorbers

Most absorbers require some intercooling between stages or trays to remove heat of absorption, and to provide internal conditions compatible with proper or required absorption. Some temperature rise (10–30°F) is usually designed into the initial conditions. The rise above this must be handled with intercoolers.

The total intercooler duty is the difference between the total heat in the rich gas and lean oil, and the total heat out of the off-gas and rich oil all at the terminal calculated or design conditions. The total duty is often divided between several coolers placed to re-cool the oil as it passes down the column. If intercoolers are not used, then the absorption cannot meet the design terminal outlet conditions, and the quantity of material absorbed will be reduced. If the intercooling is so great as to subcool, then greater absorption may be achieved, but this can be controlled by the intercooler operation.

A second approach to the same result involves the same requirements as for a balanced “heat” design; the heat of absorption of the actual components absorbed must equal the sum of the heat added to the lean

oil and to the lean gas. For hydrocarbon materials these factors can be developed by using total heats.

The relation of Hull and Raymond [32] considers heat loss through the column wall, and indicates that either the total heat of absorption, or the rich oil outlet temperature for system balance can be calculated. Thus, if a reasonable temperature balance is not obtained, the heat load for the intercoolers can be set.

(11-90)

Figure 11-26 presents actual total heats of absorption based on experimental studies [32]. As long as the hydrocarbon absorption is in the range of 80–120°F, the values read from the graph should apply.

Figure 11-26

Component heats of absorption.

Used by permission, Hall, R. J., and Raymond, K., Oil & Gas Journ. Nov. 9 (1953) thru Mar., 15 (1954).

Estimation of discharge gas temperature may be made from Figure 11-27 based on test data.

Figure 11-27

Hydrocarbon systems; overhead gas minus lean oil temperature for components absorbed in top “theoretical” tray (or top actual three trays).

Used by permission, Hall, R. J., and Raymond, K., Oil & Gas Journ. Nov. 9 (1953) thru Mar. 15 (1954).

The design of absorbers has not received the empirical design guidelines study so prevalent in distillation problems. The graph of Hutchinson [34], Figure 11-28, is convenient to examine a problem to determine preliminarily the effects of design. This curve is compatible with the conventional absorption factor graphs. The percent extraction gives a quick evaluation of the possibilities of accomplishing the desired absorption. As the number of trays in the absorber is increased, the amount of heavier material (larger A values) absorbed increases faster than the lighter components (lower A values). More heavier materials are also absorbed as the temperature of absorption is lowered. Thus, a cold lean oil has more capacity per gallon than a warm oil. This requires a study of the entire process and not just the one unit. Heat economy, oil flows, and tower costs all enter into a full evaluation of the absorber as it fits into the plant system.

Figure 11-28

Absorption equilibrium curve.

Used by permission, Hutchison, A. J. L., Petroleum Refiner, V. 29 (1950), p. 100, Gulf Pub. Co., All rights reserved.

Many designs are set up by assuming the number of theoretical trays, using the best available information for tray efficiencies and then

calculating the expected performance. A series of such studies might be made.

11.18.1. Absorption and Stripping Efficiency

Unfortunately, the efficiencies for tray and overall column operation are incomplete and nullify to a certain extent some very high quality theoretical performance design. Tray efficiencies may be estimated by Figure 10-54 or Table 11-23. Table 11-25 gives a summary of Sherwood's absorption and stripping calculation method.

Table 11-25 Summary of Sherwood Absorption a

(Source: Edmister, W. C., Hydrocarbon absorption and fractionation process de

This method is a combination of graphical and analytical using "equilibrium line" and "operating line" concepts thus illustrating diffusional driving force All compositions (X and Y) are referred to constant and known streams, such as: inlet wet gas to absorber, lean oil to absorber or from stripper, and inlet stream to stripper	Sul res we Me ma for cal
Absorption	**Strippi**
Equations: The absorption factor A has as the ratio of the lean oil to the inlet gas and term changes the lean oil to the rich oil. K is determined at an average oil temperature Estimate and calculate the amounts of each component absorbed by the above equation. Then make XY diagram for the key component	***Eq*** The lea cor ave Est the by cor
XY Diagram	***XY Diagram***

Equilibrium line is drawn with the slope at origin and values of X_R and Y_R at rich end, both being obtained from the equilibrium equation

At the lean end, , and slope of the equilibrium line = .

At the rich end, values of Y_R, and are known

Operating line is drawn with the slope and intercept on the Y axis, or the slope and trial to get proper number of steps.

Final calculations:
Repeat the analytical or graphical calculations with necessary adjustments until results of both are the same.

Eq
X_R

At

At

Op
or t

Fir
Rep
adj

*These equations which are expressions of the L/V ratio at rich end and the avera
for the key component.

Example 11-5: Determine the Number of Trays for Specified Product Absorption

A gas stream is to have the ethylene removed by absorption in a lean oil of molecular weight 160, sp. gr. 0.825. The inlet gas is at 70 psig and 100°F and the oil is at 80°F. The gas rate is 16,000,000 scf/day (60°F). Examine the tower performance for 98% ethylene recovery at 1.25 times the minimum L_o/V_N.

Feed Gas	
Component	**Mol or Volume %**

Component	Mol or Volume %
H_2	18.5
CH_4	22.3
C_2H_4	20.3
C_2H_6	0.5
C_3H_6	22.0
C_3H_8	0.7
n-C_4H_{10}	2.5
n-C_5H_{12}	13.0
	100.0

Determine the oil rate and the number of theoretical and actual trays required. (Note: this example illustrates that unreasonable results must be examined, not just accepted.)

Solution

1.

Mols ethylene in = (20.5/100) (1,756) = 360 mols/hr
Mols ethylene in outlet gas = (100 – 98/100)(360) = 7.20 mols/hr
Mols ethylene absorbed in oil = 360 − 7.2 = 352.8 mols/hr

2.

3. Minimum L/V for ethylene:

Average tower conditions for K:

Temperature = (100 + 80)/2 = 90°F
Pressure: allow 20 psi pressure drop, then top pressure would be 70 − 20 = 50 psig Average: (50 + 70)/2 = 60 psig
K (ethylene) at 60 psig and 90°F = 11.5 (from equilibrium charts).

4. Operating $(L_o/V_{N+1})_o$ = (1.25) (11.28) = 14.1
5. Operating

6. Theoretical number of trays at operating (L_o/V_{N+1}):

N = 11.29

7. Actual number of plates at operating (L_o/V_{N+1}): Efficiency of oil at 90°F, and sp. gr. of 0.825 corresponding to API of 40.
Viscosity = 0.81 centipoise
For O'Connell's efficiency correlation, Figure 10-54.

Reading curve (3), Eff. = 14%
This value is low, but agrees generally with the specific data in O'Connell's [49] report. Although Drickamer's data are not so specific for absorption, the graph of this correlation gives Eff. = 20% for the 0.81 cP. Because no better information is available, use Eff. = 15%.

Actual number of trays, $N_o = N/E_o = 11.29/0.15 = 75.26$
Use $N_o = 75$ trays
Lean oil rate = $L_o = A_i (K_i) (V_{N+1})_o = (1.227) (11.5)(1,756)$ L_o = 24,788 mols/hr

This is unreasonable, and is due to the effect of:

a. Operating pressure being too low, thus giving a high K value
b. Ethylene being a light component and difficult to absorb.
c. Temperature being too high.

8. Recalculation of steps 3 through 8.
Assume operating pressure is 700 psig, K = 1.35.
Note that this same K value could have been achieved by lowering the operating temperature to –90°F. This is also not practical from the oil standpoint or even from the economics of operating the entire system and refrigeration system at this level, unless (1) the refrigeration is available and (2) a suitable oil is available.
Min (L/V) = (1.35) (0.98) = 1.32
Operating $(L/V_{N+1})_o = (1.25) (1.32) = 1.65$
Operation $A_{io} = 1.65/1.35 = 1.222$
Theoretical number of trays at operating (L_o/V_{N+1}):

N = 11.42 trays

Reading curve 3, Eff. = 29%

Actual number of trays = 11.42/0.29 = 39.4. Say 40 trays.
Lean oil rate = (1.222) (1.35) (1,756) = 2,900 mols/hr

This is still a large quantity of oil to absorb the ethylene. Under some circumstances it might be less expensive to separate the ethylene by low-temperature fractionation.

Example 11-6: Determine Component Absorption in a Fixed-Tray Tower

An existing 40-tray tower is to be examined to determine the absorption of a rich gas of the following analysis:

Component	**Mols/hr**	**Mol wt**	**104°F K_i 176 psia**
H_2	500.0	2	59.0
CH_4	20.9	16	56.0
C_2H_4	131.5	26	8.1
CO	230.0	28	12.3
C_3H_4	3.5	40	0.07
C_4H_2	4.1	50	0.009
	890.0		

1. The key component is methyl acetylene, C_3H_4. Recovery will be based on 96.5% of this material.
The tower average temperature will be assumed = 104°F.
The operating average pressure will be = 161 psig.
2. The K_i values are tabulated for the conditions of (1), and were determined from laboratory test data for the special solvent oil being considered.
K_i = 14.7 (H_i)/176, where H_i is Henry's constant expressed as atm/mol fraction for each component.
Note that conventional K charts are only applicable to

hydrocarbon oil systems, and do not apply for any special solvents.

3. Calculate tray efficiency for C_3H_4.

Using O'Connell's correlation:

(solvent) at 104°F = 2.3 cP

Sr. gr. = 1.0

Mol. wt. solvent = 180

K = 0.07

Efficiency = 46.5% (From Figure 10-54)

Use: 45%

Actual number trays in column = 40

Theoretical number of trays based on 45% efficiency = (40) (.45) = 18

4. Using E_{ai} = 96.5% for C_3H_4. read A_i from Figure 11-24.

At n = 18

$A_e = A_i = 1.04$

$A_i = L_o/KV_{N+1}$

V_{N+1} = 890 mols/hr

L_o = (1.04) (0.07) (890) = 64.8 mols/hr

L_o/V_{N+1} = 64.8/890.0 = 0.0728

5.

Component	Inlet		
	$Y_{(N+1)i}$	Mols/hr	
H_2	0.562	500.0	0.001235

CH_4	0.0235	20.9	0.00130
C_2H_2	0.1478	131.5	0.009
CO	0.258	230.0	0.00592
C_3H_4	0.00393	3.5	1.04
C_4H_2	0.00462	4.1	8.1
	0.99985	890.0	
Component	**Mols/hr**		**Mols/hr**
	Absorbed	**X_{iR}**	**Off Gas**
H_2	0.617	0.00953	* 499.383
CH_4	0.0272	0.00042	20.8728
C_2H_2	1.185	0.0183	130.315
CO	1.36	0.021	228.64
C_3H_4	3.37	0.052	0.13
C_4H_2	4.1	0.0633	0
	10.659	0.1645	
*These are subtraction differences and do not infer that the results are this a			

Typical calculations: for hydrogen

$A_i = 0.0728/59.0 = 0.001235$

From Figure 11-24 at n = 18 theoretical number of trays, and $A_i = 0.001235$ read E_{ai}, except that in this low region some values cannot be read accurately. When A_i is considerably less than 1.0, use $E_{ai} = A_i$ (very little light material recovered), and when A_i is quite a bit larger than 1.0, use $E_{ai} = 1.0$ (heavy material mostly recovered).

Mols component absorbed/hr = $(V_{N+1})\ (Y_{(N+1)i})\ E_{ai} = (890)\ (0.562)\ (0.001235) = 0.617$

Mols component absorbed/mol lean oil = $X_{iR} = 0.0617/64.8 = 0.00953$

Mols of component in off-gas out top of absorber:
= 500.0 − 0.617 = 499.383 mols/hr
Mols component in, out top of absorber/mol inlet rich gas

6. Correcting values and recalculating

Component	Inlet		
	$Y_{(N+1)i}$	Mols/hr	
H_2	0.562	500.0	0.00144
CH_4	0.0235	20.9	0.001513
C_2H_2	0.1478	131.5	0.01048
CO	0.258	230.0	0.00689
C_3H_4	0.00393	3.5	1.21
C_4H_2	0.00462	4.1	9.43

Component	Mols/hr Absorbed	X_{iR}	Off Gas Mols/hr
H_2	0.72	0.0111	499.28
CH_4	0.0317	0.000488	20.869
C_2H_2	1.38	0.0213	130.12
CO	1.585	0.0244	228.415
C_3H_4	3.43	0.0529	0.07
C_4H_2	4.1	0.0633	0

	11.246	0.1734	878.754

Typical calculations, using hydrogen:

Mols component absorbed/hr = (0.00144) (500) = 0.72
Mols component absorbed/mol lean oil = 0.72/64.8 = 0.0111
Mols of component in off-gas out top of absorber:
= 500.0 − 0.72 = 499.28
These results do not justify recalculation for greater accuracy. Note that 98% of the C_3H_4 is absorbed instead of 96.5% as initially specified. This could be revised by reassuming a (slightly) lower oil rate, but this is not considered necessary.
The off-gas analysis Y_i represents mols gas out per mol entering rich gas.

For a new design, a study should be made of the number of trays against the required lean oil for a given absorption.

11.19. Gas Treating Troubleshooting

Troubleshooting of the glycol unit can be simplified, if the unit is appropriately instrumented, operating records, glycol analyses, process flow sheets and drawings of vessel internals are readily available [311, 312]. Interpreting the records requires knowledge of the design condition for the plant. Additionally, major changes in pertinent variables such as inlet gas flow rate, temperature or pressure require determination of a new set of optimum operating conditions. Branan [284] illustrated the corrective measures for the possible causes of common problems in glycol plant as follows.

11.19.1. High Exit Gas Dew Point

- Change in gas flow rate, temperature or pressure.
- Insufficient glycol recirculation (should be 1.5 to 3 gal TEG/lb water removed).
- Poor glycol reconcentration (exit gas dew point is 5–15°F higher than dew point in equilibrium with lean glycol concentration).

11.19.2. High Glycol Losses

First determine where loss is occurring: from contactor (i.e., down pipeline); out of still column; or leaking from pump.

- Loss from contactor
    - Inlet separator passing liquids
    - Carryover due to excessive foaming
    - Mist extractor plugged or missing
    - Lean glycol entering contactor too hot
    - Plugged trays in contactor
    - Excessive gas velocity in contactor
    - Tray spacing less than 24 in.
    - Mist pad too close to top tray

- Loss from separator
    - Glycol dumped with hydrocarbons

- Loss from still
    - Excessive stripping gas
    - Flash separator passing condensate
    - Packing in still column is broken, dirty, plugged
    - Insufficient reflux cooling at top of still
    - Too much cooling at top of still
    - Temperature at top of still is too high

- Leaks, spills, *etc.*
    - Check piping, fittings, valves, gaskets

- Check pumps, especially packings
- Pilferage

Many manufacturers will estimate maximum glycol losses to be 1 lb or 1 gas per MMscf. Improper operation can increase the actual glycol losses to 1, 10 or even 100 gallons per MMscf. One leaking pump can waste 35 gallons per day.

11.19.3. Glycol Contamination

- Carryover of oils (e.g. compressor lube oils), brine, corrosion inhibitors, well treating chemicals, sand, corrosion scales, *etc.* from inlet separator.
- Oxygen leaks into glycol storage tanks, *etc.*
- Inadequate pH control (low pH) increases corrosion.
- Overheating of glycol in reboiler due to excessive temperature or hot spots on firetube.
- Improper filtration; plugged filters; bypassing of filters.
- Improper cleaning of glycol unit; use of soaps or acids.

11.19.4. Poor Glycol Reconcentration

- Low reboiler temperature.
- Insufficient stripping gas.
- Rich glycol is leaking into lean glycol in the glycol heat exchanger or in the glycol pump.
- Overloading capacity of reboiler.
- Low fuel rate or low Btu content of fuel.
- Glycol foaming in the still column.
- Dirty and/or broken packing in the still column.
- Flooding of the still column.
- High still pressure.

11.19.5. Low Glycol Circulation – Glycol Pump

- Check pump operation. If it is a glycol powered pump, close lean glycol discharge valve; if pump continues running, it needs repair. If gas or electric, check circulation by stopping glycol discharge from contactor and timing fill rate of gauge column on chimney tray section.
- Check pump valves to see if worn or broken.
- Plugged strainer, lines, filters.
- Vapor lock in lines or pump.
- Low level in accumulator.
- Excessive packing gland leakage.
- Contactor pressure too high.

11.19.6. High Pressure Drop Across Contactor

- Excessive gas flow rate.
- Operating at pressures far below design.
- Plugged trays.
- Plugged demister pads.
- Glycol foaming.

11.19.7. High Stripping Still Temperature

- Inadequate reflux.
- Still column flooded.
- Glycol foaming.
- Carryover of light HC in rich glycol.
- Leaking reflux coil.

11.19.8. High Reboiler Pressure

- Packing in still column is broken and/or plugged with tar, dirt, *etc*.
- Restricted vent line, not sloped.
- Still column is flooded by excessive boil up rates and/or

excessive reflux cooling.

• Slug of liquid HC enters stripper, vaporizes on reaching reboiler, and blows liquid out of still.

11.19.9. Firetube Fouling/Hot Spots/Burn Out

1. Buildup of salt, dust, scales, *etc.* on firetube – check inlet separator.
2. Deposition of coke, tar formed by glycol overheating and/or hydrocarbon decomposition.
3. Liquid glycol level drops exposing firetubes. Low level shutdown.

11.19.10. High Gas Dew Points

The most obvious indication of a glycol dehydration malfunction is a high water content or dew point of the outgoing sales gas stream. In most cases, this is caused by an inadequate glycol circulation rate or by an insufficient reconcentration of the glycol. These two factors can be caused by a variety of contributing problems as illustrated in the following sections.

11.19.11. Cause – Inadequate Glycol Circulation Rate

1. Glycol powered pump. Close the dry discharge valve and see if the pump continues to run; if so, the pump needs to be repaired.
2. Gas or electric driven pump. Verify adequate circulation by shutting off the glycol discharge from the absorber and timing the fill rate in the gauge glass.
3. Pump stroking but not pumping. Check the valves to see if they are seating properly.
4. Check pump suction strainer for stoppage.
5. Open bleeder valve to eliminate “air lock”.
6. Be sure surge level is sufficiently high.

11.19.12. Low Reboiler Temperature

- Inadequate size firetube and/or burner.
- Setting temperature controller too low.
- More water in inlet gas because pressure is low or temperature high.
- Temperature controller not operating correctly.
- Carryover of water from inlet separator.
- Inaccurate reboiler thermometer.

11.19.13. Flash Separator Failure

- Check level controllers.
- Check dump valves.
- Excessive circulation.

11.20. Cause – Insufficient Reconcentration of Glycol

1. Verify the reboiler temperature with a test thermometer and be sure the temperature is in the recommended range of 350°F to 400°F for triethylene glycol. The temperature can be raised closer to 400°F, if needed, to remove more water from the glycol.
2. Check the glycol-to-glycol heat exchanger in the accumulator for leakage of wet, rich glycol into the dry, lean glycol.
3. Check the stripping gas, if applicable, to be sure there is a positive flow of gas. Be sure steam is not back flowing into the accumulator from the reboiler.

11.20.1. Cause – Operating Conditions Different from Design

1. Increase the absorber pressure, if needed. This may require the installation of a back pressure valve.
2. Reduce the gas temperature, if possible.
3. Increase the glycol circulation rate, if possible.

11.20.2. Cause – Low Gas Flow Rates

1. If the absorber has access manways, blank off a portion of the bubble caps or valve trays.
2. Add external cooling to the lean glycol and balance the glycol circulation rate for the low gas rate.
3. Change to a smaller absorber designed for the lower rate, if needed.

11.20.3. High Glycol Loss

Cause – Foaming

Foaming is usually caused by contamination of glycol with salt, hydrocarbons, dust, mud and corrosion inhibitors. Remove the source of contamination with effective gas cleaning ahead of the absorber, improved solids filtration, and carbon purification.

Cause – Excessive Velocity in the Absorber

1. Decrease the gas rate.
2. Increase the pressure on the absorber, if possible.

Cause – Trays Plugged With Mud, Sludge, and Other Contaminants

If the pressure drop across the absorber exceeds about 15 psi, the trays may be dirty and/or plugged. Plugged trays and/or downcomers usually prevent the easy flow of gas and glycol through the absorber. If the absorber has access handholes, manual cleaning can be helpful. Chemical cleaning is recommended if handholes are not available.

11.20.4. Cause – Loss of Glycol Out of Still Column

1. Be sure the stripping gas valve is open and the accumulator is vented to the atmosphere.
2. Be sure the reboiler is not overloaded with free water entering with the gas stream.
3. Be sure excessive hydrocarbons are kept out of the reboiler.

4. Replace the tower packings in the still column, if fouled or powdered.

If the outlet gas temperature from the absorber exceeds the inlet gas temperature by more than 20 to 30°F, the lean glycol entering the top of the absorber may be too hot. This could indicate a heat exchanger problem or an excessive glycol circulation rate.

If a glycol pump has been operating in a clean system, no major service will probably be needed for several years. Only a yearly replacement of packing is usually required. Normally the pump will not stop pumping unless some internal part has been bent, worn, or broken, a foreign object has fouled the pump, or the system has lost its glycol. A pump which has been running without glycol for some time should be checked before returning to normal service. The pump will probably need at least new “O” rings. The cylinders and piston rods may also have been scored from the “dry run”.

The following are typical symptoms and causes for a Kimray pump operation.

Symptoms

1. The pump will not operate.
2. The pump will start and run until the glycol returns from the absorber. The pump then stops or slows appreciably and will not run at its rated speed.
3. The pump operates until the system temperature is normal and then the pump speeds up and cavitates.
4. The pump lopes or pumps on one side only.
5. Pump stops and leaks excessive gas from the wet glycol discharge.
6. Erratic pump speed. Pump changes speed very few minutes.
7. Broken pilot piston.

Causes

1. One or more of the flow lines to the pump are completely blocked or the system pressure is too low for standard pumps.
2. The wet glycol discharge line to the reboiler is restricted. A pressure gauge installed on the line will show the restriction immediately.
3. The suction line is too small and an increase in temperature and pumping rate cavitates the pump.
4. A leaky check valve, a foreign object lodged under a check valve, or a leaky piston.
5. Look for metal chips or shavings under the pump D-slides.
6. Traps in the wet glycol power piping send alternate slugs of glycol and gas to the pump.
7. Insufficient glycol to the Main Piston D-slide ports. Elevate the control valve end of the pump to correct.

The glycol pH should be controlled to avoid equipment corrosion. Some possible causes of a low, acidic pH are:

1. Thermal decomposition, caused by an excessive reboiler temperature (over 404°F), deposits on the fire tube, or a poor reboiler design.
2. Glycol oxidation, caused by getting oxygen into the glycol with the incoming gas; it can be sucked in through a leaking pump or through unblanketed glycol storage tanks.
3. Acid gas pickup from the incoming gas stream.

Salt accumulations and other deposits on the fire tube can sometimes be detected by smelling the vapors from the still vent. A “burned” odor emitted from these vapors usually indicates this type of thermal degradation. Another detection method is to observe the glycol color.

It will darken quickly if the glycol degrades. These detection methods may prevent a fire tube failure.

In general, maintenance and production records together with the used lean and rich glycol analyses can be very useful to the troubleshooter. A history of filter element, carbon, tower packing, and fire tube changeouts can be shown and the frequency of pump repairs and chemical cleaning jobs are most helpful to the troubleshooter, who can quickly eliminate and thus prevent costly problems.

11.21. Sour Water Stripping Process

Sour water streams containing hydrogen sulfide (H_2S), other organic sulfur compounds, ammonia (NH_3), and oil are collected from various refinery process units and are combined in a feed surge tank. Liquid hydrocarbons are decanted from the water and returned to the recovery oil tank. The Sour Water Stripper (SWS) removes hydrogen sulfide and ammonia (H_2S/NH_3) from the sour water using a stripper tower having a steam heated reboiler.

A sour water stripper either uses a steam fired reboiler as a heat source or direct application of steam (usually low quality and low pressure). There are no fired heaters associated with the sour water stripper. The purpose of this unit is to drive as much H_2S and NH_3 overhead in the stripper as possible. The sizing of the unit is essential, since its capacity must equal or exceed the normal production rates of sour water from various sources in the refinery. Generally, refiners find their strippers undersized because of a lack of provision of handling large amounts of sour water, which can result from upset conditions (e.g. startup and shut down). There is often a large quantity of sour water waiting to be processed in the stripper, and in the light of the increasing importance of environmental restrictions and regulations by governments (e.g. U.S. EPA), the sour water stripper

unit plays an important role in mitigating the overall pollution of refiners.

Figure 11-29 shows the process flow diagram where the feed sour water is preheated by exchange with the stripper bottoms stream. The reboiler is heated with low pressure steam to generate a vapor stream flowing up the stripper column. Vaporization of water strips H_2S and NH_3 from the downcoming sour water. Overheads vapors are cooled by an overhead condenser. The condensed water reflux is returned to the top tray in the stripper tower. The overhead, noncondensable materials, primarily H_2S and NH_3 are routed to the sulfur recovery unit as feed. The stripped water is reused at the crude desalters and at process units requiring wash water (e.g. for ammonia removal). Any remaining stripped water is routed to the washwater treatment plant.

Figure 11-29

Process flow diagram of a sour water stripping unit.

A Case Study

80,000 gal/day of sour water at 100°F and 40.0 psia goes through a feed/effluent exchanger where it recovers heat from the tower bottoms stream (stripper bottoms). The new stream (stripper feed) at 200°F enters on tray 2 of an 8 stage distillation column with a reboiler and a total reflux condenser. A quality specification of 10 ppm wt. ammonia on the tower bottoms (stripper bottoms) is specified. The tower bottoms (stripper bottoms) exchange heat with the incoming feed and exists as effluent.

The sour water feed compositions are:

Component	Mass fraction
Hydrogen sulfide (H_2S)	0.007
Ammonia (NH_3)	0.005
Water (H_2O)	0.988

Solution

UniSim simulation software Sour-Water-Unit USC design R 370 is used to simulate the sour water at 100°F and 40.0 psia entering the tube side of the heat exchanger with stripper bottoms entering the shell side of the exchanger. The stripper feed stream exits the outlet of tube side and enters the sour water stripping column at the second stage of an 8 plate distillation column, which is used to strip the sour components from the feed stream. The liquid leaving the bottom of the column heats the incoming feed stream in the heat exchanger.

There are two basic steps in setting up the simulation process:

1. Setup. This case study uses the Sour Peng-Robinson package and the following components: H_2S, NH_3 and H_2O.
2. Steady State Simulation. The case consists of an 8 stage stripper, used to separate H_2S and NH_3 and uses a heat exchanger to minimize heat loss.

The specifications are as follows:

Feed Stream

Specify the feed stream as shown.

In this cell	Enter

Temperature	100°F
Pressure	40 psia
Std Ideal Liq Vol Flow	80,000 barrel/day
Comp. Mass Frac (H_2S)	0.0070
Comp. Mass Frac. (NH_3)	0.0050
Comp. Mass Frac. (H_2O)	0.9880

Operations

1. Install and specify the heat exchanger as:

Tab [Page]	In this cell	Enter
Design [Connections]	Tube Side Inlet	Sour H_2O Feed
	Tube Side Outlet	Stripper Feed
	Shell Side Inlet	Stripper Bottoms
	Shell Side Outlet	Effluent
Design [Parameters]	Heat Exchanger Model	Exchanger Design (Weighted)
	Tube Side Delta P	10 psi
	Shell Side Delta P	10 psi
Worksheet [Conditions]	Temperature (stripper Feed)	200°F

2. Install a distillation column. The column will have both a reboiler and an overhead condenser.
3. Define the column configuration as shown below.

	In this cell	Enter
Connections	No. of stages	8
	Inlet Stream	Stripper Feed
	Inlet stage	2
	Condenser Type	Full Reflux
	Ohd Vapor	Off Gas
	Bottoms Liquid	Stripper Bottoms
	Reboiler Energy Stream	Q-Reb
	Condenser Energy Stream	Q-Cond
Pressure Profile	Condenser Pressure	28.7 psia
	Reboiler Pressure	32.7 psia

4. In the column property view, click the **Design** tab, then select the **Monitor** page.
In the present configuration, the column has two degrees of freedom, the two specifications used will be a quality specification and a reflux ratio.
5. Modify the existing specification as shown below

Column [SW Stripper]		
Tab [Page]	**Variable Spec**	**Modify**
Design [Specs]	Ovhd Vap Rate	Active = uncheck
	Reflux Ratio	Active = checked Spec Value = 10 Molar

6. Add a Component Fraction specification, and enter the

following information in the Comp Frac Spec view

Tab	In this cell	Enter
Parameters	Name	NH_3 Mass Frac. (Reboiler)
	Stage	Reboiler
	Spec Value	0.000010
	Component	Ammonia
Summary	Active	Checked
	Reflux Ratio Spec Value	Active 10 Molar

7. Click the Parameters tab, then select the Solver page. Change the Fixed Damping Factor to 0.4.
A damping factor will speed up tower convergence and reduce the effects of any oscillations in the calculations (the default value is 1.0).
8. Run the column to calculate the values by clicking the **Run** button.
The results of the simulation are shown in Appendix L.

Glossary

Absorption: A separation process involving the transfer of a substance from a gaseous phase to a liquid phase through the phase boundary.

Acid Gases: Impurities in a gas stream usually consisting of H_2S, CO_2, COS, RSH, and SO_2. Most common in natural gas are of H_2S, CO_2, and COS.

Adsorption: The process by which gaseous components adhere to solids because of their molecular attraction to the solid surface.

Alkanolamine: An organic nitrogen bearing compound related to ammonia by having at least one of its hydrogen atoms substituted with at least one linear or branched alkanol groups where only

one or two could also be substituted with a linear or branched alkyl group (i.e. methyldiethanolamine MDEA). The number of hydrogen atoms substituted by alkanol or alkyl groups at the amino site determines whether the alkanolamine is primary, secondary or tertiary.

Amine treating: Contacting a gas or light hydrocarbon liquid with an aqueous solution of an amine compound to remove the H_2S and CO_2.

API Gravity: A method for reporting the density of petroleum streams. It is defined as [(145.1/Sg) – 131.5] where Sg is the specific gravity relative to water. API gravity is reported at a reference temperature of 60°F (15.9°C).

Aromatics: Derivatives of benzene, a six carbon atom ring structure with three double bonds. Aromatics may have multiple combined benzene rings, with side chains consisting of other chemical structures attached to the carbon atoms in the ring structure.

Asphalt: Heavy tar-like residue from distillation of some types of crude oil. Asphalt components are high molecular weight derivatives of aromatic compounds. Not all asphalt materials are suitable for use as building agents in road pavement.

Asphaltenes: Highly condensed masses of high molecular weight aromatic compounds. They exist in petroleum residue as the center of colloidal particles or micelles. The asphaltenes are kept in solution by an outer ring of aromatic compounds of lower molecular weight. They can precipitate when the continuous nature of the surrounding ring of aromatics is broken down by cracking processes.

ASTM: American Society of Testing and Materials. Nearly all of the refinery products tests have been standardized by ASTM.

Atmospheric Residuum: The heaviest material from the distillation of crude oil in a crude distillation column operating at a positive pressure.

Azeotrope: A constant boiling point mixture for which the vapor and liquid are of identical composition. Components of azeotropes cannot be separated with conventional distillation.

Boiling point: Typically refers to the temperature at which a component or mixture of components start to vaporize at a given pressure. When used in petroleum refining it is usually synonymous with the normal boiling point (boiling point at one atmosphere).

Boiling range: The lowest through highest boiling temperatures for a petroleum stream when distilled. Boiling ranges are often reported on a TBP (true boiling point) basis, *i.e.* as normal boiling points.

Bubble point: The temperature at which a component or mixture of components begins to vaporize at a given pressure.

Carbon number: The number of carbon atoms in one molecule of a given hydrocarbon.

Characterization factor: A calculated factor used to correlate properties for petroleum streams. It is a measure of the paraffinicity of the stream and is defined as ([MABP∗∗1/3]/Sg) where MABP = mean average boiling temperature in degrees (°R) and Sg = specific gravity at 60°F (15.9°C) relative to water.

Catalytic cracking: Process in which an oil feed is reacted in a hot catalyst bed to produce heating oils, gasoline, and light petroleum liquids and gases. The products are stripped from the catalyst and the catalyst is burned in a regeneration step to remove coke deposits resulting from the cracking reactions.

Catalytic polymerization (cat.poly): Process in which propylene and or butylene components are chemically joined to produce gasoline. A phosphoric acid catalyst is usually employed in the process.

Catalytic reforming (reforming): Process where low octane naphthas are chemically changed into high octane gasoline and light gas products over a platinum catalyst.

Continuous Catalytic Reforming (CCR) process: is where the catalyst is circulated through the reactors and a regeneration step, analogous to catalytic cracking processes.

Cetane number: A number calculated from the API gravity and the D86 fifty percent distilled for a petroleum stock. It is used to rate the performance of a fuel in diesel engines.

Cloud Point: The temperature at which the first crystals form when a petroleum stock is cooled in a test apparatus.

Coke: A carbonaceous deposit formed by the thermal decomposition of petroleum.

Coking: Refining process in which petroleum oil is heated destructively such that the heaviest materials are converted to coke. There are two processes: delayed coking and fluid coking, with delayed coking being the most widely used.

Condensation: Reaction in which aromatic ring structures combine to form ring structures larger than the reactants.

Conversion: The fraction of a feed stock converted to gasoline and lighter components.

Cracking: The process by which the composition of petroleum is changed by heat induced reactions. Cracking may or may not take place in the presence of a catalyst.

Critical point: The temperature and pressure at which a component or mixture of components enter a dense phase, being neither liquid nor vapor.

Critical Pressure: The vapor pressure at the critical temperature.

Critical Temperature: The temperature above which a component cannot be liquefied. For mixtures, the temperature above which all of the mixture cannot be liquefied.

Crude oil: A complex mixture of hydrocarbons ranging from methane (CH_4) to compounds boiling higher than 2000°F (1094°C). It is a natural resource and is pumped from underground locations.

Cut point temperature, cut points: Arbitrarily defined

temperatures on the laboratory boiling point curve for a petroleum stock. The boiling point curve most commonly used to define cut points is the true boiling point curve (TBP) at one atmosphere of pressure.

Cut point ranges: A series of cut point temperatures are defined for a petroleum stock. The cut point ranges are the temperature differences between adjacent cut point temperatures. When developing petroleum pseudo-components for a petroleum stock, cut point ranges must be defined which include the total boiling point range of the stock.

Cycle oil, cycle stock: An oil stock containing hydrogen deficient compounds that was produced in a thermal or catalytic cracking operation.

Cyclo-olefin: Unsaturated ring structure with one or two double bonds in the ring.

D86: ASTM laboratory distillation method conducted at atmospheric pressure for streams boiling below approximately 700°F (371°C). The D86 is the most commonly used laboratory distillation for petroleum stocks.

D1160: ASTM laboratory distillation method for high boiling streams. The D1160 is performed under vacuum conditions with 10 mm Hg being the most common pressure used for the test. D1160 data are normally reported on a 760 mm Hg basis.

D2287: ASTM simulated distillation method for high boiling streams. The D2887 has an upper limit of 1000°F (538°C) and the temperatures are reported vs. weight per cent distilled. A normal paraffin standard is used to convert the chromatographic results to a boiling point curve.

D3710: ASTM simulated distillation method for gasoline and light naphthas. D3710 data are reported on a volume basis.

Debutanizer: A column that removes n-butane (nC_4H_{10}) and lighter components in the top product.

Deethanizer: A column that removes ethane (C_2H_6) and lighter

components in the top product.

Delayed coking: A refinery process in which heavy oil is heated and then allowed to thermally crack and deposit coke in a large drum that provides delay time for the coking reaction to occur. The coke is removed from the drum with a hydraulic drill that is lowered into the drum from a derrick that sits on top of the drum.

Dehydration: a process used to remove water from natural gas and natural gas liquids (NGLs), and is required to prevent formation of hydrates, corrosion and condensation of free water in processing and transportation.

Dehydrogenation: A chemical reaction in which a compound loses bonded hydrogen.

Deisobutanizer: A column that removes isobutene (iC_4H_8) and lighter components in the top product.

Demethanizer: A column that removes methane (CH_4) and lighter components in the top product.

Depentanizer: A column that removes n-pentane (C_5H_{12}) and lighter components in the top product.

Desalting: A process that removes chlorides and other inorganic salts from crude oil by injecting water and applying an electrostatic field to force the salt into the aqueous phase.

Desuperheating zone: A section of a distilling column where a superheated vapor is cooled and some liquid is condensed. FCC main fractionators have a desuperheating zone.

Dew point: The temperature and pressure at which the first drop of liquid will condense for a component or mixture of components.

Diesel fuel: A fuel produced for diesel engines with typical ASTM D86 boiling point ranges of 450–675°F (233–358°C).

Distributed component: A component that appears in both the top and bottom products from a distillation column separation zone.

Downcomer: A device to direct the liquid from a distillation

column at a location above the bottom tray and below the top tray. Draws may be vapor or liquid phase.

Draw, side draw: A product stream withdrawn from a distillation column at a location above the bottom tray and below the top tray. Draws may be vapor or liquid phase.

Endothermic: a process that requires heat input.

End point (final boiling point): The highest boiling point recorded for a laboratory distillation. Usually, there is some residual material in the laboratory still, and the end point is not the highest boiling point material in the mixture being distilled.

Energy balance: Summation of the energy entering a process and the summation of the energy leaving a process. They must be equal for a steady state process.

Equilibrium K value (K value): The ratio of the mole fraction in the vapor phase divided by the mole fraction in the liquid phase for a component in the equilibrium state. Each K value corresponds to a given temperature, pressure and mixture composition.

Entrainment: A non-equilibrium process by which liquids are mechanically carried into a vapor that is leaving a process vessel or contacting device.

Exothermic: Exothermic processes release heat as a byproduct.

Flash calculation: Determination of the compositions and quantities of liquid and vapor that coexist in a mixture under equilibrium conditions.

Flash point: The temperature at which the vapors above a heated hydrocarbon mixture will ignite when contacted by an open flame. It is determined in a laboratory.

Flash zone: The section of a distillation column containing the column feed nozzle(s). The column feed separates or “flashes” into liquid and vapor as it expands through the feed nozzle(s) and enters the column.

Flooding: occurs in a distillation column when flows of liquid

and/or vapor greater than the column can process are encountered. It can be caused by liquid backing up in the column, vapor blowing through the column and lifting the liquid off the trays, by mechanical restrictions or damage to the internals in the column, *etc.* All columns are designed to handle about 80% of the flow before flooding occurs.

Fluid catalytic cracking (FCC): The most commonly used catalytic cracking process. The catalyst is a fine powder that is designed to form a fluidized bed in the reactor and regenerator.

Fluid coking: involves the feed being preheated and sprayed into a reactor where it contacts a hot fluidized bed of coke returning from a burner vessel. The hot oil products are stripped from the coke which is circulated back to the burner vessel. Coke not returned to the reactor from the burner vessel is withdrawn as a coke product.

Fractionator: Another term for a distillation column.

Free energy of formation: The change in free energy when a compound is formed from its elements with each substance in its standard state at 77°F (25°C). The heat of reaction at 25°C may be calculated by subtracting the sum of the free energies of formation of the reactants from the sum of the free energies of formation of the products.

Free water: Condensed water that exists as a separate liquid phase. Most refinery distillation columns are designed such that free water will not be present, since it can result in column upsets and promote corrosion of the metal in the column.

Fixed bed reactor: A reactor in which the catalyst is loaded into an immovable bed. The reactants enter the top of the bed, and the products exit from the bottom of the bed. The process must be taken off line, and hot gases circulated through the catalyst bed to burn off coke deposits and restore the catalyst activity.

Gap: is usually based on ASTM D86 distillation temperatures and is defined as the 5% distilled temperature of a distillation

column product minus the 95% distilled temperature of the next higher product in the column. When the difference is positive, the difference is called a gap. When the difference is negative, the difference is sometimes called an overlap. The gap or overlap is a measure of the sharpness of the separation between adjacent products in a distillation column.

Gas oil: is a term used for petroleum stocks with boiling ranges between 650–1100°F (344–594°C). Uncracked gas oils are produced by distilling crude oil in crude and vacuum columns. Cracked gas oils are produced in refinery reaction processes, such as thermal and catalytic cracking, coking, visbreaking and hydrocracking.

Gas treating: Amine treating of light gases to remove impurities such as H_2S and CO_2. Molecular sieves are also used to concentrate hydrogen streams by removing inerts and light hydrocarbon contaminants.

Gasoline: A name for petroleum stocks that are used as fuel for internal combustion engines. Retail gasoline is a blend of several refinery gasolines, which must meet specifications for octane, Reid vapor pressure, distillation boiling range, sulfur content, *etc.* Additives such as ethers or alcohols are used to improve the octane rating for the blended product.

Grain: A unit of mass where one pound is equivalent to 7000 grains and a specification of 0.25 grain of H_2S per 100 scf is equivalent to an H_2S concentration of 4.0 ppmv.

Gravity: The specific gravity of a stream, often expressed in API Gravity by petroleum refiners. The basis is always the density of water.

Gravity curve: The gravity of the material distilled from a petroleum stock in a laboratory still. The gravity curve is plotted against the percent distilled for the stock. Gravity curves are most commonly reported for true boiling point distillation data.

Gross Heating Value: is the total energy transferred as heat in

an ideal combustion reaction at a standard temperature and pressure in which all water formed appears as liquid. The gross heating is an ideal gas property in a hypothetical state (the water cannot all condense to liquid because some of the water would saturate the CO_2 in the products).

HSTD: High temperature simulated distilled. Laboratory test designed for petroleum stocks boiling up to 1382°F (750°C).

Heat of reaction: The heat released or heat absorbed when a chemical reaction takes place. The heat of reaction may be calculated from the free energies of formation for the reacting components and the resultant products at the standard temperature of 77°F (25°C).

Heavy cycle oil (HCO): Gas oil produced in a FCC operation that boils in the approximate TBP range of 400–1000°F (205–538°C). Heavy cycle oil is not withdrawn as a product, but rather it is recycled to the reactor for further cracking in order to improve the overall conversion of the process.

Heavy Key: A distributed component in a distillation section that is recovered in the (bottoms) heavy product.

Hydrocarbon: Any organic compound that consists of hydrocarbon and carbon atoms.

Hydrocracking: A refining process where a petroleum stream is cracked over a catalyst in a high hydrogen atmosphere. The high H_2 partial pressure immediately saturates any unsaturated compound produced by the cracking reactions.

Hydrotreating: a process where cracking is minimized. The hydrogen consumed in chemical reactions is that required to convert sulfur and nitrogen compounds to H_2S and NH_3, plus hydrogen that replaces carbon/carbon double bonds in saturation reactions. Some hydrogen is lost through solubility in the liquid product(s).

Initial boiling point: the boiling temperature in a laboratory still at which the first drop of distilled liquid is condensed. The initial

boiling point may be higher than the boiling point of light components in the sample, since these are not condensed by the apparatus.

Hypothetical State: is defined as a fluid in a state that cannot actually exist, *e.g.* methane as a liquid at 60°F and 14.696 psia. Methane cannot be in its liquid phase at this temperature and pressure, but such a state when defined, can be used in calculations.

Isomerization: A reaction process where the molecular structure of a compound is rearranged to a different isomer form, but the chemical formula remains unchanged.

Jet fuel: A kerosene material of typical ASTM D86 boiling point range 400–550°F (205–288°C) used as fuel for commercial jet aircraft.

Joule Thompson expansion: The pressure of a mixture is reduced with no heat transfer to or from the surroundings. The pressure decrease typically results in a temperature decrease with the exception for systems that consist largely of hydrogen gas.

Kerosene: A material from distillation of crude oil that boils in the approximate ASTM D86 range of 400–550°F (205–288°C).

Key components: In a conventional distillation column with two products, two components or groups of components define the separation. Both components must be distributed to the top and bottoms products. The light key appears in the bottoms product in a small significant quantity and the heavy key appears in the top product in a small significant quantity.

Lean oil: The absorption oil entering the top tray of an absorption column.

Light cycle oil (LCO): Gas oil produced in a catalytic cracking operation that boils in the approximate ASTM D86 range of 400–695°F (205–369°C).

Light key: A distributed component in a distillation section that is recovered in the top light product, with a small specified

amount leaving the bottoms product.

Liquefied petroleum gas (LPG): Liquefied petroleum gas, typically mixtures of C_3 and C_4 hydrocarbons.

Make up stream: A feed to a process to replace a component that reacts or is otherwise depleted in a process.

Mass balance: Summation of the mass entering a process and the summation of the mass leaving a process. They must be equal for a steady state process.

Mixed phase: More than one phase; usually vapor and liquid phases are present.

Molecular sieve: A separation process that usually works by gaseous diffusion. A membrane is selected through which the compounds being removed or purified can pass while the remaining compounds in the stream being processed cannot pass.

Methyl tertiary butyl ether (MTBE): A gasoline additive used to increase octane rating. MTBE is produced by reacting methanol with isobutylene.

Naphtha: A term that is applied to low boiling mixtures of hydrocarbons, with typical TBP boiling ranges between 150–450°F (66–233°C). Light and heavy naphthas are produced in the distillation of crude oils. Cracked naphthas are also produced by many of the refinery reaction processes.

Naphthenes: Hydrocarbons of the cyclane family, sometimes called cycloalkanes. Naphthenes have no double bonds and are saturated ring structures with the general formula C_nH_{2n}, where C = carbon atoms and H = Hydrogen atoms.

Net Heating Value: is the total energy transferred as heat in an ideal combustion reaction at a standard temperature and pressure in which all water formed appears as vapor. The net heating is an ideal gas property in a hypothetical state (the water cannot all remain vapor because, after the water saturates the CO_2 in the products, the rest would condense).

Octane: A test used to measure the suitability of a gasoline as

motor fuel. The octane test determines the knocking characteristics of a gasoline in a standard test engine relative to a standard 2-2-4 trimethyl pentane (224 TMP). 224 TMP is assigned an octane number of 100. There are two octane tests. One is designated the research octane (F-1), and the second as the motor octane (F-2). Motor octane is determined in an engine more representative of actual operating conditions for automobiles and is lower than research octane for any gasoline stock.

Off gas: The gas leaving a reflux drum or top of tray of an absorber column.

Olefins: Hydrocarbon of the alkene family. Olefins have two carbon atoms in the molecular structure linked by a double bond to satisfy the absence of two hydrogen atoms that are present in the corresponding paraffin. The hydrogen deficiency is called unsaturation. The general formula for olefins is C_nH_{2n}, where C = Carbon atoms and H = Hydrogen atoms.

Overhead: Usually refers to the vapor leaving the top tray of a distillation column. For an absorber column, the overhead and the top product are the same.

Paraffins: Hydrocarbons of the alkane family. Paraffins are saturated compounds, that is, hydrogen atoms are appropriately attached to the carbon atoms such that the carbon atoms have only single bonds in the molecular structure. The general formula for paraffins is C_nH_{2n+2}, where C = carbon atoms and H = Hydrogen atoms.

Partial pressure: In a gaseous mixture, the pressure contribution for a particular component of the mixture.

Phase Envelope: The boundaries of an area on the P-T diagram for the material which encloses the region where both vapor and liquid coexist.

Physical Solvent: A liquid capable of absorbing selected gas components by solubility alone without associated chemical

reactions.

Polymerization: A reaction in which like molecules are joined together to form dimer and trimer compounds etc of the reactant(s). This most often occurs with olefinic compounds.

Pour point: A sample tube of petroleum oil is chilled in the pour point test. The pour point is defined as the temperature at which the sample will still pour (move) when the sample tube is tipped. The pour temperature is typically about 5°F (2.8°C) lower than the cloud point.

ppmv: A volume concentration of a species in a bulk fluid measured in parts per million.

Preheat, preheat train: Heat exchanger or a net work of heat exchangers in which the feed to a process (usually a distillation column) is heated by recovering heat from products being cooled.

Pumparound: A liquid side draw from a distillation column that is pumped, cooled and returned to a higher location in the column. Pumparound recover useable heat that would be lost at the condenser. They also lower the vapor flow in a column and reduce the required column diameter for vapor loaded columns such as crude and vacuum columns.

Purge: A stream that is removed from a recycle process to prevent buildup of one or more compounds in the process streams.

Quality: The weight fraction of vapor in a vapor-liquid mixture.

Quench stream: A cooled stream that is used to cool another stream by direct contact, *eg.* Hydrogen quench streams are used to quench the hot effluents from hydrocracker reactors.

Quench zone: A section of a distillation column where a hot stream, usually vapor, is cooled by direct contact with a stream that has been cooled, usually a liquid.

Rating calculations: Calculations in which a unit operation such as a column, heat exchanger, pump, *etc.* is checked for capacity

restrictions.

Reactive distillation: A distillation column in which there is a section designed for chemical reaction, usually containing a catalyst bed. Some methyl acetate, MTBE and TAME processes use a reactive distillation column in place of a second reactor prior to the product stream.

Rectification zone: the section of a distillation column in which heavy components flow down the column by contact with a liquid reflux stream. In a conventional distillation columns, this is the section of the column from the tray above the feed tray to the top tray.

Recycle: A process stream that is returned to an upstream operation.

Reduced Pressure: The ratio of the absolute pressure to the critical pressure.

Reduced Temperature: The ratio of the absolute temperature to the critical temperature.

Reformate: The gasoline produced in a catalytic reforming operation.

Reflux: Condensed liquid that is returned to the top tray of a distillation column. Reflux helps rectify the mixture being distilled by washing heavy components down the column.

Reflux drum: A drum that receives the outlet from the overhead condenser for a distillation column. The liquid and vapor portions are separated in the reflux drum.

Reid vapor pressure (RVP): An ASTM test method to determine the vapor pressure of a light petroleum stream. The Reid vapor pressure is very nearly equal to the true vapor pressure for gasoline streams. There is also a Reid vapor pressure test for crude oil.

Residence Time: The period of time in which a process stream will be contained within a certain volume or piece of equipment, in seconds.

Rich oil: The absorption oil leaving the bottom tray of an absorption column. The rich oil contains the absorbed light components.

Saturated compounds: Hydrocarbons in which there are no double bonds between carbon atoms. Saturated compounds contain the maximum possible number of hydrogen atoms.

Saybolt Seconds Universal (SSU): The viscosity of a petroleum stock as measured by an ASTM laboratory test. The test value is the time in seconds for a measured sample of the stock to flow through an orifice at a given temperature. SSU is normally measured at temperatures of 100 and 210°F (38.1 and 99.2°C)

Secondary absorber: The second absorber in a FCC gas plant is usually the last unit operation in the gas recovery plant and is known as the sponge absorber.

Selective Treating: Preferential removal of one acid gas component, leaving at least some of the other acid gas components in the treated stream.

Separation zone: A section of a distillation column in which a separation between two products occurs. Components which are found in both products are said to be distributed components.

Separator: is referred to as a drum, in which residence time is provided for a mixture of liquid and vapor to separate into liquid and vapor streams. It is also called a flash drum. The liquid and vapor leaving the separator are in phase equilibrium.

Side heater (reboiler): A heat input to a distillation column that is located above the bottom tray of the column.

Side stripper: A small auxiliary column that receives a liquid draw product from a main distillation column, for stripping out of light components. Light components are stripped by a stripping steam, or reboiling and returned to the main column. Liquid products are sometimes stripped in side strippers to raise the flash point.

Simulated distillation: A new laboratory method in which a

petroleum stream is separated into fractions with gas phase chromatography. Carbon disulfide (CS_2) is used as the carrying agent to dissolve the petroleum stream. The component fractions elute from the chromatographic column in a time sequence that is related to their boiling temperatures. Temperatures are assigned to the fractions based on the chromatographic separation of a normal paraffin standard mixture. The simulated distillation approaches a true boiling point distillation, and is reported on a mass basis for streams heavier than gasoline. Aromatic compounds elute from the column faster than paraffins of similar boiling points. Hence, simulated distillations must be corrected for aromatic content when stocks contain significant quantities of these compounds.

Slurry: It is bottom stream from FCC main fractionators, and is referred to as slurry because it contains suspended catalyst particles.

Sour gas: A light gas stream that contains acid gases, *e.g.* sulfur compounds (e.g. hydrogen sulfide), ammonia compounds and/or carbon dioxide.

Splitter: A distillation column that separates a feed into light and heavy products.

Specific gravity: by definition is the ratio of gas density (at the temperature and pressure of the gas) to the density of dry air (at the air temperature and pressure).

Sponge oil: Absorption oil that is fed to the top tray of the sponge (secondary) absorber.

Stabilizer: A distillation column that removes light components from a liquid product. This term is used to describe debutanizer columns that remove C4 hydrocarbons from gasoline to in order control the vapor pressure.

Steady state: A process where the mass and energy flowing both into and out of the process are balanced.

Stoichiometric: Applied to reactors in which the reactants and

products are defined in terms of the molar quantities reacting, *eg.* . The stoichiometric coefficients are −3, −2, and 2 for H_2, N_2 and NH_3 respectively.

Straight run: A product that has been distilled from crude oil but has not been through a process in which its composition has been chemically altered.

Stripper column: A loose designation applied to a distillation column in which light components are stripped from a heavier liquid product.

Stripping steam: Steam that is injected into the bottom of a side stripping column or used to strip oil from catalyst in a FCC operation.

Stripping zone: The section of the column in which light components are stripped from a heavier liquid product. In conventional distillation columns, this is the portion of the column from the reboiler to the feed tray.

Sweet gas: Gas which has no more than the maximum sulfur content defined by the specifications for the sales gas from a plant (i.e. a gas stream from which the sulfur compounds have been removed).

Tail gas: the residue gas from an absorption column. In FCC gas plants, it is referred as the gas products from the secondary absorber column.

Tertiary amyl methyl ether (TAME): used to enhance the octane rating of a motor gasoline pool. TAME is produced by reacting the isoamylenes produced in FCC processes with methanol.

Tetra ethyl lead (TEL): A compound added to gasoline to increase the octane rating. TEL has been superseded by other octane enhancers and is no longer used by refiners for motor gasoline.

Theoretical tray: refers to a vapor/liquid contact device (e.g. distillation column) in which the liquid and vapor leaving the device are in perfect phase equilibrium. There are also perfect energy and mass balances for a theoretical tray.

Thermal cracking: The first cracking process, in which the oil was cracked by heating only. Thermal cracking produces lower octane gasoline than catalytic processes.

Top product: For columns with condensers, the liquid and/or vapor streams from the reflux drum that exit the process.

Topped crude oil: The bottom product from a crude distillation column.

Tray: A liquid/vapor contact device in a distillation column.

Threshold Limit Value: The amount of a contaminant to which a person can have repeated exposure for an eight hour day without adverse effects.

True boiling point distillation (TBP): A laboratory test in which a petroleum oil is distilled in a column having at least 15 theoretical trays and a reflux ratio of 5.0. The distillate is continually removed and further analyzed. The separation corresponds somewhat to a component by component separation and is a good measure of the true composition for the sample being distilled. As the temperatures in the still increase, the pressure of the still is lowered to suppress thermal cracking of the sample. The minimum pressure for most TBP stills is about 38 mm Hg. This allows distillation of petroleum components boiling up to about 900–950°F (483–510°C) at a pressure of 1 atm.

Two phase: A mixture consisting of one vapor in equilibrium with one homogeneous liquid phase.

Undistributed component: A component in a distillation column separation zone that is totally recovered in only one of the products.

Unsaturated compounds: Hydrocarbon compounds in which some of the carbon atoms have multiple bonds with other carbon

atoms because of lack of hydrogen atoms to satisfy the carbon atom valencies.

Vacuum distillation: A distillation column that operates at subatmospheric pressure. Vacuum distillation allows the further distillation of heavy feed stocks at reduced temperatures and to minimize cracking reactions.

Vacuum residuum: The heaviest product from a vacuum distillation column.

Visbreaking: A process in which a heavy oil is thermally cracked just enough to lower or break the viscosity. A small quantity of gas oil and lighter products is formed in the process.

Wash zone: A section in a column where the column vapor is washed of entrained heavy materials by contact with a cooler injected liquid. A section of packed material is often used to promote good mixing of the liquid and vapor in the wash zone. All vacuum distillation columns have wash zones to remove heavy residual material that is carried up the column from the flash zone. If washing is inadequate, the heavy residual material forms petroleum coke and plugs the column above the flash zone.

Nomenclature

A = actual desiccant bed area, ft^2

A_{DES} = calculated area based on desiccant volume, ft^2

Cycle = drying cycle, hr

D_{BED} = calculated bed diameter, ft

E = joint efficiency

F = loss factor for non-steady state heating

Flow = gas flow rate,

h = additional shell height for desiccant supports and distributor, ft

H_1, H_2 = enthalpy of regeneration gas before and after

regeneration heater, Btu/lb

H_{GAS} = total regeneration gas requirement, lb

H_{REG} = heat added to regeneration gas to regenerate desiccant bed

H_{TOT} = heat required to regenerate dryer (ideal), Btu

HT = heater temperature, °F

H_2O load = total water adsorbed, lb

H_2O Cont = water content of gas, lb/10^6 std ft^3

L = desiccant bed length, ft

MW = molecular weight of regeneration gas

P = dryer operating pressure, psig

PICKUP = useful design capacity of desiccant, %.

Q = actual gas flow rate at flowing conditions, ft^3/min

R = radius of desiccant bed, ft

S = allowable stress, lb/in^2

SCF = standard cube feet of regeneration gas

t = dryer shell thickness, inch

T = gas flowing temperature, ºR

T_{BED} = temperature of desiccant bed, °F

V_{DES} = desiccant volume, ft^3

V_{GAS} = superficial gas velocity, ft/min.

W_D = desiccant weight, lb

WS = steel weight of shell and heads, lb

W_{VES} = vessel weight, lb

Z = compressibility factor

ρ_D = desiccant density, lb/ft^3

Nomenclature For Part 2, Absorption, Stripping and Adsorption

(Special notations, all others same as for Distillation Performance Nomenclature, Part 1)

A′ = Edmister's effective absorption factor

A" = Outside surface area of absorber, ft2
A = Absorption factor, average
A" = Effective absorptive factor
A_{Bi} = Absorption factor for each component at conditions of bottom tray
A_{Ti} = Absorption factor for each component at conditions of top tray
Cp = cp = Specific heat, Btu/lb (°F)
E_a = Absorption Efficiency, or fraction absorbed
E_o = Overall tray efficiency, fraction
E, = Stripping efficiency, or fraction stripped
f_{ai} = Fraction of v_{n+1i} absorbed by the liquid
f_{si} = Fraction of l_{oi} stripped out of the liquid
G_{mi} = Mols individual components stripped per hour
ΔH = Total heat of absorption of absorbed components, thousand Btu/day
K = Equilibrium constant, equals y/x, at average tower conditions
L_{M+1} = Mols/hour rich oil entering stripper
L_N = Liquid leaving bottom absorber tray, mols/hr
L_o = Mols/hr lean oil entering absorber, or leaving stripper
L_{oi} = Molar flow of component “i” in entering liquid to absorber
M = Number theoretical stages in stripper
m = M (see above)
N = Number theoretical stages in absorber
n = N (see above)
S, S_e = Stripping factor, average and effective, respectively
S′ = Edmister's effective stripping factor
T_i = Tray i, temperature, °F
T_N = Bottom tray temperature, °F
T_{N+1} = Inlet rich gas temperature, °F
T_o = Lean oil temperature, °F

U = Overall heat transfer coefficient between absorber outside surface and atmosphere, BUI/(ft^2) (°F) (hr); usual value = 3.0

V_1 = Mols/hr lean gas leaving absorber

V_i = Gas leaving tray i, mols/hr

V_{i+1} = Gas leaving tray, i + 1, mols/hr

V_N = Vapor leaving bottom absorber tray, mols/hr

V_{N+1} = Mols/hr rich gas entering absorber

V_o = Mols/hr stripping medium (steam or gas) entering stripper

V_{ii} = Molar gas flowrate of component "i" leaving plate 1 in absorber

V_{n+1i} = Molar gas flowrate of component "i" in entering gas to absorber

W = Rate of flow, thousand lb/day

X = Number mols absorbed component or stripped per mol lean oil entering column

X_1 = Number mols liquid phase component in equilibrium with Y_1

X_{iR} = Mols of a component in liquid absorbed per mol of lean oil entering column

L_{Xi} = Total mols of all liquid phase components absorbed per mol of lean oil (omitting lean oil present in liquid phase, considered = 1.0)

X_{M+1} = Number liquid phase mols of component entering stripper per mol of lean oil

X_{oi} = Number liquid phase mols of component entering absorber with lean oil per mol of lean oil

Y_1 = Number vapor phase mols of component leaving top plate of absorber per mol rich gas entering absorber

Y_i = Mols component in vapor phase from tray "i"/mol rich gas entering absorber

ΣY_i = Total mols of all vapor phase components stripped per mol

of stripping medium

Y_{N+1} = Number vapor phase mols of component entering absorber per mol rich gas entering

Y_o^* = Number vapor phase mols of component in equilibrium with lean oil per mol of rich gas entering

Subscripts

1, 2, etc = Components in a system

Amb = Ambient

Avg, A_v = Arithmetic average

DG = Discharge gas

e = Effective

i = Individual components in mixture

IG = Intake gas

Key = Key component

L = Lean concentration end of column

L_o = Lean oil

Min = Minimum condition

o = Operating condition

R = Rich concentration end of column

R_o = Rich oil

S = Absorbed components

Chapter 12. Enhanced distillation types

Several distillation methods have been developed for close boiling or low relative volatility systems (e.g. $\alpha < 1.10$) and for those that exhibit azeotropic characteristics. These techniques rely on the differences between the vapor and liquid compositions, as does ordinary distillation, but an additional mechanism is used to alter the behavior of key components. These techniques are as illustrated by Stichlmair, Fair and Bravo [287] and Seader and Henley [255], and are listed below.

12.1. Homogeneous azeotropic distillation

This is a technique for separating a mixture by adding an entrainer, which then forms a homogeneous minimum or maximum boiling azeotrope with one or more feed components. It should be recalled that azeotropic systems show a minimum in the T–x–y diagram when the deviations from Raoult's law are positive (Figure 10-25a), and a maximum in the T–x–y diagram when the deviations from Raoult's law are negative (Figure 10-26a). If under these two conditions, a single liquid phase is in equilibrium with the vapor phase, the azeotrope is homogeneous.

The entrainer is added near the top of the column to the feed, or near the bottom of the column depending on whether the azeotrope is removed from the top or bottom.

Figure 12-1 shows the flowsheet of an isopropyl alcohol and water mixture. Here, azeotropic distillation is used to separate a high purity product; this cannot be achieved by ordinary fractional distillation.

Figure 12-1

Azeotropic distillation process for separation of isopropyl alcohol and water.

During normal distillation of isopropyl alcohol and water mixture, the liquid composition reaches 91% isopropyl alcohol and 9% water, which is an azeotrope. This is a constant boiling mixture, so no further changes in composition will occur. To achieve a purer isopropyl alcohol product, the azeotrope must be broken by adding ethyl ether. This forms an other azeotrope, or constant boiling mixture with water, but having a lower boiling point. This means that, during distillation, the ether-water mixture will be preferentially removed as overhead product and 99.9% isopropyl alcohol is left as a bottoms product. After the overhead product leaves the distillation tower, it goes to a recovery station where the ether is separated from the ether–water mixture and recycled to the top of the azeotropic distillation column. When ether or some entrainer is used to break an azeotrope and form a new one, large quantities of heat are consumed. This is because the entrainer used to break the azeotrope vaporizes and carries the unwanted component overhead. Thus the entrainer used should have a low heat of vaporization. Figure 12-2a and Figure 12-2b show T–x–y and VLE diagrams of a 2-propanol-water mixture at 1 atmosphere exhibiting a minimum boiling azeotrope.

Figure 12-2a

Phase diagram for the mixture 2-propanol and water at 1 atm.

Figure 12-2b

VLE Diagram for the mixture 2-propanol and water at 1 atm.

12.2. Separation of Minimum Boiling Homogeneous Azeotropes

The boiling point diagram for an azeotropic mixture whose boiling point changes with measure is shown two different pressures is shown in Figure 12-3. Such an azeotrope may be separated by employing two columns as shown in Figure 12-4. The feed is introduced to the second column at the higher pressure P_2. The bottom product of this distillation contains relatively pure component B, and the top product consists of the minimum boiling azeotrope $x_{D,A} = 0.8$ and $x_{D,B} = 0.2$. This distillate is fed to the first column which is operated at the total pressure of P_1. This column produces a bottom product that is pure A. The top product is the azeotropic in its composition at P_1 ($x_{D,A} = 0.6$ and $x_{D,B} = 0.4$) and thus is added to the feed for the second column that is maintained at pressure P_2.

Figure 12-3

Boiling point diagram for a minimum boiling homogeneous azeotropic which is

sensitive to a change in the total pressure.

Figure 12-4

Use of two columns to separate a minimum-boiling homogeneous azeotrope which is sensitive to a change in the total pressure.

12.3. Separation of Maximum Boiling Homogeneous Azeotropes

Separation of these azeotropes may be effected in a similar way to that described for minimum boiling azeotropes. Figure 12-5 shows the boiling point diagram at pressures P_1 and P_2. Separation may also be effected by the use of two columns, as illustrated in Figure 12-6. The azeotrope may be separated by feeding the column operated at P_1 with the initial mixture, and removing pure B as distillate product and the azeotrope of composition obtained at this pressure as the bottoms product. This is then fed to the column operated at P_2 and pure A is obtained as distillate, with the azeotrope of composition obtained at P_2 obtained as the bottoms product. This can then be combined with the first feed to the plant.

Figure 12-5

Boiling point diagram for a maximum-boiling homogeneous azeotrope which is sensitive to a change in the total pressure.

Figure 12-6

Use of two columns to separate a maximum-boiling homogeneous azeotrope which is pressure sensitive.

12.4. Heterogeneous Azeotropic Distillation

Where the components of the mixture are not miscible in all proportions with each other, and form two seperate phases at some ratios the azeotrope is heterogeneous. For heterogeneous azeotropes, the vapor-phase composition is equal to the overall composition of two or more liquid phases. The boiling point diagram for a mixture of components A and B is shown in Figure 12-7.

Figure 12-7

Boiling point diagram for a heterogeneous minimum-boiling azeotrope.

If a minimum boiling azeotrope is formed by the entrainer. The azeotrope is distilled from the feed as a mixed vapor, which then condenses into two liquid phases in the overhead condensing system, one of which is returned to the column reflux, while the other is sent

to another separation unit or is a product.

If the feed is in the twophase region *i.e.* it is present as two seperate liquid mixtures - of A dissolved in B, and B dissolved in A, the two phases may be separated and fed to each of the two columns. The feed is introduced to the separator as shown in Figure 12-8 which operates at the temperature T_{sep}. Phase I is then fed to column I and phase II to column 2. The bottom products of column 1 and 2 are pure B and A respectively. The distillate compositions are approximately those of the azeotrope, and consequently can be fed to the separator.

Figure 12-8

Use of a two-column system to separate a heterogeneous minimum-boiling azeotrope.

One of the mechanisms for identifying entrainers is hydrogen bonding. Ewell *et al.* [319] have classified liquids into five groups according to their hydrogen-bonding characteristics. For example, the following are considered strong and weak in their hydrogen bonding:

The hydrogen bonding grouping provides insight for screening suitable entrainers in the development of a feasible azeotropic distillation process. Selected entrainers are then tested experimentally for their quantitative effect on vapor liquid equilibrium (VLE). Ewell *et al.* [319] developed workable guidelines to identify chemical

classes suitable for azeotropic distillation entrainers. These guidelines are reviewed by Berg [320]. The basis for the classification of Ewell and co-workers' guidelines is the idea that liquid properties are related to the degree of bonding between molecules. The hydrogen bond is the most important of the bonds and serves as the most important criterion.

Class I. Liquids are capable of forming three-dimensional networks of strong hydrogen bonds, and include water, glycol, glycerol, amino acids, hydroxyl acids and amides.

Class II. Other liquids composed of molecules containing both active hydrogen atoms and donor atoms of oxygen, nitrogen and fluorine. These include alcohols, acids, primary phenols, oximes, primary and secondary amines, ammonia, hydrogen fluoride, hydrogen cyanide, nitro compounds with alpha-hydrogen atoms, *etc*.

Class III. Liquids containing donor atoms but no active hydrogen atoms. These include ethers, aldehydes, esters, ketones, tertiary amines, nitro compounds, and nitriles without alpha-hydrogen atoms.

Class IV. Liquids composed of molecules with active hydrogen atoms but no donor atoms. These include molecules with two or three chlorine atoms on the same carbon atom as a hydrogen atom, or one chlorine atom on the same carbon and one or more chlorine atom on adjacent carbon atoms ($CHCl_3$, CH_2Cl_2, CH_3CHCl_2, $CH_2Cl{-}CHCl_2$, etc.).

Class V. All other liquids – those without hydrogen-bond forming capability. These include hydrocarbons, carbon disulfide, mercaptans, halohydrocarbons not in Class IV, and nonmetallic elements.

In mixtures, liquids from these classes may form hydrogen bonds or

may break them and as such introduce deviations from Raoult's law. We should remember that positive deviations from Raoult's law may lead to minimum boiling azeotropes and negative deviations from Raoult's law to maximum boiling azeotropes. Table 12-1 is useful in identifying classes of compounds for screening tests, although not all classes are effective nor will all members of a given class be useful.

Table 12-1 Deviations of class combinations from Raoult's law (Source: M. Van Winkle, Distillation, Mc.Graw-Hill, New York, 1967).

Classes	Deviations	H bonding
I + V	Always +	H bonds broken only
II + V	I + V frequently showing limited solubility	
III + IV	Always -	H bonds formed only
I + IV	Always +, I + IV limited solubility	H bonds both formed and broken (dissociation common)
II + IV		
I + I		
I + II	Usually +	H bonds both broken and formed
I + III	Some – giving maximum boiling azeotropes	
II + II		
II + III		
III + III		
III + V	Quasi-ideal always	No H bonds involved
IV + IV	+ or ideal	
IV + V	Minimum azeotropes	
V + V	Minimum azeotropes (if any)	

Close – boiling compounds

Azeotropes are useful in separating such mixtures when:

1. The entrainer forms a binary minimum boiling azeotropes with only one component in a binary mixture.
2. The entrainer forms binary minimum boiling azeotropes with both compounds, but one azeotrope has a sufficiently lower boiling point than the other.
3. The entrainer forms a ternary minimum boiling azeotrope with a boiling point sufficiently below that of any binary azeotrope. The ratio of the original feed components in the ternary must be different from the ratio in feed, and the ternary (preferably heterogeneous) must be separable in some way.

12.5. Minimum Boiling Azeotropes

1. The entrainer forms a binary minimum-boiling azeotrope which has a boiling point sufficiently below that of the original azeotrope.
2. The entrainer forms a ternary minimum-boiling azeotrope with a sufficiently low boiling point and with a different ratio of the original components from that in the original azeotrope.

Ewell *et al.* [319] classified maximum boiling azeotropes as:

1. Water + strong acids: HCl, HNO_3, HBr. *etc.*
2. Water + associated liquid: formic acid, hydrazine
3. Donor (III) liquids + nonassociated liquids having active H atoms (IV): acetone + chloroform
4. Organic acids + amines
5. Phenols + amines
6. Organic acids + donor liquids containing O_2: formic acid + diethyl ketone
7. Phenols + donor liquids containing O_2: phenol + methyl hexylketone
8. Phenols + alcohols: phenol + n – octanol.

12.6. Salt Distillation

The technique involves dissolving a suitable inorganic salt in the top reflux of the process, such that the relative volatilities of the components of the mixture are altered. Since the salt is nonvolatile, it remains in the liquid phase as it flows down the column. See Seader and Henley [255] for a detailed process description of this method.

12.7. Pressure Swing Distillation

When a binary azeotrope disappears at some pressure or changes composition by 5 mol % or more over a moderate range of pressure, a technique is used where two columns in sequence are operated at different pressures. Knapp and Doherty [321] have provided 36 pressure sensitive binary azeotropes based upon the compilation of Horsley [322].

12.8. Distillation with Vapor Recompression (VRC)

This technique uses heat pumps to reduce the energy requirement of distillation. In this method, no water steam is required in the reboiler. The vapor from the top of the column is passed into a compressor which increases its pressure. The discharged vapor is then fed into the reboiler where, through its condensation, vapor is generated in the column. Proper selection of column internals is essential; various type of structured packings such as Mellapak Plus provide very low pressure drop and high capacity. This is important because the efficiency of the heat pump depends on the pressure difference between column top and bottom.

Distillation with VRC can be both environmentally friendly and economical where the temperature difference between the top and bottom is low, and where the heat duty in the reboiler is high. Meszaros [323] has listed many industrial processes where distillation with VRC has been successfully employed. Figure 12-9 and Figure

12-10 provide a comparison between conventional distillation and with vapor recompressions, and Table 12-2 and Table 12-3 show the advantage and applications of the technique. Application of this technique in industrial processes is described by Meili [324], Meszaros and Meili [325] and Meszaros and Hangi [326].

Figure 12-9

Typical scheme of conventional distillation

(adapted from Meszaros, I., Environmental-friendly and economical distillation with heat pump [290]).

Figure 12-10

Typical scheme of distillation with vapor recompression (VRC)

(adapted from Meszaros, I., Environmental-friendly and economical distillation with heat pump [290]).

Table 12-2 Incentive for application of vapor recompression for distillation (Source: Meszaros, I., Environmental-friendly and economical distillation with heat pump [290]).

Advantages	Explanation
No steam consumption	Apart from steam turbine driven compressor the process does not need steam. In this way boiler house extension can be avoided which would be otherwise necessary for conventional process

	otherwise necessary for conventional process
Decisively lower cooling water requirement	Because the energy is recovered and utilized in the same process the net energy requirement is lower. This involves lower cooling water requirement which is needed to set the energy balance
The operation is more independent of cooling water disturbances	This is a consequence of much lower cooling water consumption
Less expensive safety auxiliary systems	In emergency the fire case determines the required blow down capacity which is normally much smaller
The heat pumping technology contributes to an active environment protection and in this way ISO 14000 conforms	A technology striving for the possible lowest energy consumption and satisfies the specification of ISO 14000
The heat pumping technology pays less CO_2 tax	An economic incentive is realized by introduction of separate tax connected to the CO_2 emission. Due to the smaller primary energy consumption heat pumping technology pays less CO_2 tax
Higher safety due to moderate pressure increase at cooling water failure	This is a consequence of much lower cooling water consumption
Better quality of temperature-sensitive product	Lower temperature levels in the reboiler reduces the decomposition and polymerization
Safer operation due to less fouling in the reboiler	Lower temperature levels in the reboiler reduces the decomposition and polymerization
Financial support can be awarded for production using technologies of low energy consumption	Government support energy saving efforts

Table 12-3 Incentive for application of vapor recompression for distillation (Source: Meszaros, I., Environmental-friendly and economical distillation with heat pump [290]).

Separations with vapor recompression in operation	A few other possible applications
1-Butene	Acetic acid
Chlorobenzene	Butanol
Dichlorobenzene	Chloronitrobenzene
Dimethyl formamide	Cumene
Ethanol	Cyclohexane
Ethylene glycol	Methanol
Hydrogen peroxide	Methylphenol (Cresol)
Isobutane	Phenol
Isopropanol	Xylene
Propylene	Further applications and many others
Silane	
Styrene	

12.9. Extractive Distillation

This is a technique that employs a large amount of relatively high boiling solvent that alters the liquid phase activity coefficients of the mixture. This allows the relative volatility of the key components to be come more favorable to separation. This process employs an agent to modify the relative volatility ( ) between the key components without forming an azeotrope. It is usually used to improve the separation between close boiling components where ordinary distillation is economically unfeasible. The method, like azeotropic distillation, involves highly nonideal phase equilibria and the addition of an agent to shift the phase equilibria of the system components to more favourable values. However, no azeotropes are formed and the agent added is essentially nonvolatile. It is customary

to call the added agent the solvent rather than the entrainer.

The technique is used to separate key components with a relative volatility ( ) < 1.1. The solvent boils so far above the system components that the formation of new azeotropes is impossible. Because of its low volatility, the solvent always leaves the extractive column with the bottom product. In other to maintain the solvent throughout the column, it must be introduced with or above the fresh feed.

The choice of a suitable solvent for the process is easier than the selection of an entrainer for azeotropic distillation, being based on having a similar structure to the heavy key component. This means that the high-boiling solvent will decrease the volatility of this component. Since the light component has a different structure to the solvent, its volatility is less affected, so increasing the difference in volatility between the two components. The design of azeotropic or extractive distillation systems requires significant preliminary investigations such as:

- Developing and finding necessary data
- Preliminary screening
- Choosing the solvent/entrainer
- Small-scale testing
- Computer simulation.

The requirements for choosing the ideal solvent or entrainer include the following characteristics:

1. It must affect the vapor-liquid equilibrium relations of the key components (relative volatilities,).
2. It should have a low latent heat.
3. It must be nonreactive with the other components in the feed

mixture.
4. It must be easily separable from the components with which it leaves the column.
5. It should be noncorrosive and nontoxic.
6. It should be inexpensive.
7. It should remain soluble in the feed components and should not lead to the formation of two phases.
8. It must be compatible with the overall process and product specifications.
9. It should be environmentally friendly.

No single solvent or solvent mixture would satisfy all the above criteria, so compromises must be reached. Tassios [327] has presented techniques for rapid screening of solvents for extractive distillation using VLPC units.

Figure 12-11 shows an extractive distillation process for the separation of methylcyclohexane and toluene, with phenol as the solvent.

Figure 12-11

Extractive distillation process for separation of methylcyclohexane and toluene.

12.10. Residue Curve Maps

A residue curve map (RCM) is a collection curves showing how the composition of a ternary mixture changes when distilled. Different residue curves result from different initial compositions for a given

pressure. Residue curve maps contain the same information as phase diagrams, but represent information in a way more pertinent to understanding how to separate a mixture by distillation.

If the direction of the curves is taken to be from starting composition to end composition, then the curves will start from the light (lowest boiling) pure components in a mixture, move towards the intermediate boiling component, and terminate, at the composition, a pure component that boils at the highest temperature. Residue curves begin and end at nodes. The lowest temperature nodes are referred to as *unstable nodes* (UN), as all trajectories leave from them, while the highest temperature points in the region are referred to as *stable nodes* (SN), since all trajectories end there. The point that the trajectories approach from one direction and end in a different direction (i.e. the boiling point of an intermediate boiling component) is referred to as the *saddle point* (S). A residue curve that separates the composition space into different distillation regimes is referred to as a distillation boundary. It connects two fixed points on nodes to a saddle. The distillation boundaries divide the separation space into separation regions. The shape of the distillation boundary plays an essential role in assessing how to separate the mixtures. Figure 12-12 shows a typical residue curve map with arrows showing the directions from the lowest temperature component towards the highest. The residue curve map thus enables one to view the distillation boundaries and distillation regions in the component space. Table 12-4 shows the characteristics of a residue curve map.

Figure 12-12

Residue curve for the ternary mixture benzene, cyclohexane and propylene

glycol at 1 atm.

Table 12-4 Characteristics of a residue curve map

1. The residue curves always point in the direction of increasing temperature 2. Azeotropes give rise to boundaries in the composition space, signaling different qualitative behavior depending on where one is in the composition space 3. Pure components and azeotropes are fixed points in the map 4. Distillation boundaries separate distillation regions, which cannot be crossed by simple stage-by-stage distillation. A distillation region can be considered to be a collection of residue curves which all start at the same unstable nodes and end at the same stable node. The distillation boundary is then the dividing line between families of residue curves 5. Residue curves are equivalent to the composition profile in a packed distillation tower at fixed pressure and infinite reflux

Consider the simple Rayleigh batch or differential distillation as shown in Figure 10-58. For any component of a ternary mixture, a material balance for its vaporization from the liquid in the still, assuming that the liquid is perfectly mixed and at its bubble point temperature is:

(12-1)

where

= mole fraction of component i in W moles of perfectly mixed liquid residue in the still

= mole fraction of component i in the vapor (instantaneous distillate) in equilibrium with x_i.

Since W changes (decreases) with time t, W and t can be combined into a single variable. From Doherty and Perkins [328], if we let this variable be . Therefore;

(12-2)

Combining Equations (12-1) and (12-2) to eliminate ;

(12-3)

Let the initial condition be x = 0 and W = W_o at t = 0. Then the solution to Equation 12-3 for at time t is:

(12-4)

Since W(t) decreases with time, (which is dimensionless, and known as wrapped time) must increase with time. Therefore, for the ternary mixture, the simple distillation process can be modeled by a set of differential algebraic equations (DAEs),

assuming that a second liquid phase does not form:

(12-5)

(12-6)

(12-7)

and the bubble point temperature equation:

(12-8)

where, in the general case,

Therefore, the system consists of seven equations with nine variables: and . If the pressure is fixed, the next seven variables can be calculated from Equations (12-5)(12-6)(12-7) and (12-8) as a function of from a specified initial condition. The calculations can proceed in either the forward or backward direction of . The results, when plotted on a triangular graph are called a residue curve because the plot follows, the liquid-residue composition in the still over time. Figure 12-13 illustrates three possible forms of residue curve patterns and Figure 12-14 shows a selection of residue curve

maps that are favorable for homogeneous azeotropic distillation. Figure 12-15 shows the residue curve map of an isopropanol, n-propanol and water mixture, highlighting its key features.

Figure 12-13

Schematics of residue curve maps when there is one binary minimum boiling azeotrope

(source: Doherty, M.F., and M. Malone, Conceptual Design of Distillation System, McGraw-Hill, New York, [2001]).

Figure 12-14

Residue curve patterns (a) near pure-component vertices; (b) near binary azeotropes; (c) near tenary azeotropes

(source: Doherty, M.F., and G.A. Caldarola, IEC Fundam., 24, 477 [1985]).

Figure 12-15

Sketch of a residue curve map: isopropanol, n-propanol, water

(source: Dimian, A.C., and C.S. Bildea, Chemical Process Design–Computer Aided Case Studies, Wiley-VCH [2008]).

Figure 12-16 shows residue curve maps for nonideal mixtures with azeotropes, Figure 12-16 a shows one distillation field for a acetone-heptane-benzene plot and Figure 12-16 b shows one distillation boundary that links the higher-boiler with the low-boiler azeotrope in an acetone-chloroform-toluene plot, which results in two distillation regions. Similar behavior is shown in Figure 12-16 c, which has two azeotropes. The acetone-methanol-chloroform plot in Figure 12-16 d has four azeotropes (three binaries and one ternary) and four distillation regions.

Figure 12-16

Residue curve maps for nonideal ternary systems involving azeotropes
(source: Dimian, A.C., and C.S. Bildea, Chemical Process Design–Computer Aided Case Studies, Wiley-VCH [2008]).

The analysis of the structure of the residue curve maps has been extensively reviewed by Doherty and Perkins [328], Van Dongen and Doherty [329] and Phan and Doherty [330]. The basic result from these analyses is that only nodes and saddles can occur in residue curve maps. For homogeneous azeotropes, the nodes may be stable (high boiling azeotropes), unstable (low boiling azeotropes) or saddles (medium boiling azeotropes). However, heterogeneous azeotropes can be either unstable nodes or saddles, but they cannot be stable nodes (i.e. they cannot be high boiling point azeotropes).

Research studies have been carried out to determine the relationship between the number of nodes (stable and unstable) and saddle points for a ternary residue plot. The topological arguments are based on an equation of the form:

(12-9)

where

N_i = number of nodes (stable and unstable) involving i species

S_i = number of saddles involving i species.

Many different residue curve maps are possible when azeotropes are present. Ternary mixtures containing only one azeotrope may exhibit six possible residue curve maps that differ by the binary pair forming the azeotrope and by whether the azeotrope is minimum or maximum boiling. Although a simple distillation process has no practical use as a method for separating such mixtures, this technique has useful application for:

- Predicting the order and content of the cuts in batch distillation
- testing the consistency of experimental azeotropic data
- checking whether the given mixture is separable by distillation
- identifying suitable entrainers/solvents
- predicting attainable product compositions
- designing the corresponding distillation column.

By identifying the limiting separation that may be achieved by a distillation process, residue curve maps are essential in synthesizing separation sequences that combine distillation with other methods. Steps for developing residue curves are illustrated by Doherty and Malone [262] and by Seader and Henley [255]. Recently, Luyben [331] employed Aspen Plus simulation software to plot residue curve for a methyl acetate, methanol and water ternary system, and Excel software for a residue curve map for the ethanol-methanol-water system has been developed by Lee Partin (SIMTOOLS, Case 3) *. The files, with instructions, can be accessed from the companion web

site of this book. In general, these plots have been shown to provide useful insight into the design of complex systems. Table 12-5 gives general heuristics for sequencing the separation of liquid mixtures. The following are comments on Table 12-5 [332].

1. Remove troublesome species as early as possible, with the priority on lights.
2. "Obtain a high-purity component as the top product in the final distillation" is the "golden rule" in industry. For recycle streams sent to reactors, the purity should be not necessarily high, but should be constant. Obviously, harmful impurities for the catalyst should be removed.
3. Distillation remains the most reliable separation technique in present use, and is the most efficient in a large number of cases. Therefore, it should be tried first and other separation methods should be rated against it.
4. The partition of a mixture in to zeotropic and azeotropic submixtures tremendously simplifies the separation problem. Some components may be distributed between the two subsystems, in which case these are coupled by recycles and bridge separations.
5. Difficult zeotropic separations should not appear in the middle of a separation sequence, because this can create a bottleneck, but in the last place before azeotropic separations. A low relative volatility separation demands a large number of stages and a high reflux ratio. This has high energy costs is an essential issue, and other separation methods should be explored first, such as adsorption, molecular sieving and crystallization.
6. The separation of azeotropic mixtures is complex. Homogeneous azeotropic distillation is appealing but finding a suitable entrainer is difficult. In a number of cases, such as breaking water/alcohol azeotropes, the heterogeneous azeotropic distillation gives much better results. Extractive distillation is a good alternative when a high-boiling solvent may be found.

Often combining distillation with other separation techniques is recommended.

7. Systematically removing the most plentiful component in intermediate mixtures will drastically reduce the cost of separation, both in capital costs and energy consumption. These heuristics typically lead to a direct sequence.

8. When the components are evenly distributed, a 50/50 split is recommended since it leads to a drastic reduction of the number of downstream separations.

*The author is grateful to Mr. Lee Partin for providing this software of residue curve map for homogeneous ethanol-methanol-water system.

Table 12.5 General heuristics for separation sequencing of liquid mixtures (Source: Fair, J. R., Design Aspects of Reactive Distillation, Chem. Eng. P 158, Oct. [1998]).

1. Remove first corrosive, hazardous, fouling, reactive and any troublesome components 2. Deliver high-purity products as top distillate. The same is valid for reactants sent to reactors sensitive to impurities 3. When separation by distillation is feasible, prefer it in a first attempt

4. Isolate geotropic and azeotropic mixtures

5. Perform difficult zeotropic separations later, but before azeotropic separations Examine other options, such as extractive distillation, L–L extraction, crystallization, adsorption, or molecular sieves

6. Examine the separation of azeotropic mixtures lasts

7. Remove the components in order of decreasing percentage of the feed. This operation will reduce the cost of the next separation
8. Favor 50/50 splits

12.11. Reactive (Catalytic) Distillation

Reactive distillation is a process where a chemical reaction occurs in the still, as it is the integration of reaction and distillation into a single unit where the products can be removed to drive a reversible reaction to completion. Separation of the product from the reaction mixture does not require a separate distillation unit, thus saving energy (e.g. for heating) and materials. This technique is especially favorable for systems where equilibrium reactions occur, such as esterification and ester hydrolysis reactions, but undesirable when one of the desired products decomposes. Conversion to product can be increased far beyond what is expected by the equilibrium due to the continuous removal of reaction products from the reaction zone, thus reducing capital and investment costs. The following heuristics illustrate the potential attractiveness of reactive distillation if [333]:

- the conventional process contains a reactor that require cooling, provided that placing the volume needed for the reactor does not increase the column size
- a product is subject to a degradation reaction, but can be removed from the reaction phase by evaporation
- the distillation has an unwanted azeotrope (a mixture of two or more pure chemicals in a ratio that cannot be changed simply through distillation) that can be broken by reactive distillation.

Research on many aspects of reactive distillation such as modeling

and simulation, process synthesis, column design, nonlinear dynamics, and control is in progress. The suitability of reactive distillation for a particular reaction depends on various factors, such as volatilities of reactants and products, as well as the feasibility of the required reaction and distillation temperature. The conditions in the reactive column are sub-optimal, both as a chemical reactor and as a distillation column since the reactive column combines these units. The introduction of an in situ separation process in the reaction zone results in complex interactions between vapor–liquid equilibrium, mass transfer, and diffusion and chemical kinetics, which pose a great challenge for the design and synthesis of these systems. Side reactors where a separate column feeds a reactor and vice versa is more suitable for some reactions where optimal conditions of reaction and distillation greatly differ. Thus, reactive distillation technique is not feasible for all processes.

When a catalyst is used, the combined operation is referred to as catalytic distillation, and is found to be suited to chemical reactions limited by equilibrium constraints, since one or more of the products of the reaction is (are) continuously separated from the reactants. The determination of feasible distillation sequences, design of the columns, and optimization of the column operating conditions for ordinary distillation of multicomponent mixtures, as illustrated earlier in the book, can be tedious but is relatively straight forward. In contrast, determining and optimizing operating conditions for this type of operation is much more difficult. This is because of liquid solution nonidealities and the difficulty of specifying feasible separations. However, this aspect of work has been solved by the graphical techniques described by Partin [334], Peterson and Partin [335], Doherty and Caldarola [336] and Matsuyama and Nishimura [337].

This section reviews the reactive distillation technique, since other

enhanced distillation types and the design of tenary systems are well illustrated by Doherty and Malone [262], Siirola and Barnicki [338] and Stichlmair and Fair [263]. A case study of the production of methyl acetate from the reaction between methanol and acetic acid, using simulation software (Honeywell UniSim®) is presented.

Distillation accompanied by chemical reaction has been shown to be more economical in many catalyzed reactions, such as esterification, etherification, hydrolysis, alkylation, nitration and amidation reactions. Reactive distillation was studied extensively as early as in the 1920s by Backhaus [339], and has been the subject of various reviews as illustrated by Doherty and Buzad [340]. One of the advantages of this process is that both reaction and separation occur simultaneously in the same vessel, and the products can be removed to drive a reversible reaction to completion.

Reactive distillation is employed whenever a liquid phase reaction is carried out with a large excess of one reactant, where conventional processes incur high costs due to the recycle of the excess reactant. In such a case, reactive distillation can be employed under stoichiometric feed conditions, so eliminating recycle costs. The large excess of are reactant is used to increase the conversion of the limiting reactant. The continuous separation of products from reactants that occurs in reactive distillation causes the reaction to proceed in the forward direction by Le Chatelier's principle (e.g. manufacture of esters such as methylacetate from methanol and acetic acid). There are other applications where the chemical reaction has the effect of removing contaminants or overcoming azeotropes in the mixture and thus simplifying the phase behavior, as in the production of methyl tertiary butyl ether (MTBE) and methyl acetate.

The design of a reactive distillation process is more complex than conventional distillation. A suitable catalyst is chosen for either

homogeneous or heterogeneous reaction. The chosen catalyst often influences the column internal specifications (e.g. plates or packings), catalyst baskets, the liquid holdup and its distribution throughout the column. Additionally, the reaction zone must be optimally placed within the column. There are cases where the reaction takes place along the whole of the column and some where only a section of the column is active. The nonreactive zones can provide purification by ordinary distillation or can be used to break an azeotrope with an entrainer.

The advantages of reactive distillation are the separation and recycling of unwanted reactants, elimination of complicated product recovery, processes, with resultant in savings in equipment and energy costs, improved reactant conversion, selectivity and mass transfer.

12.11.1. Column Design

Effective design of a reactive distillation process requires knowledge of reaction kinetics, equilibrium conversion rates, phase behavior, hydraulic characteristics and economics relevant to the process under consideration.

Operating parameters such as process variables, temperature and pressure must be compatible with the reaction conditions since both reaction and separation are carried out simultaneously.

Solid catalysts for heterogeneous reaction must be placed in the column so as not to create an excessive pressure drop their use should not necessitate a large diameter column. Only one liquid phase should be present, since a second can introduce problems in the hydraulic characteristics of the column. If two liquids phases make contact with the solid catalyst, it might segregate or the rates of component diffusion from each phase to the catalyst surface might be difficult to predict.

Commercial simulation software (Aspen, Hysys, Chemcad, ProII, Honeywell UniSim, etc.) can now determine concentration profiles throughout the column. Stage rate and phase equilibrium models that combine both reaction and separation are used, and can also be applied to reaction mixtures that exhibit nonideal behavior.

Fair [341] provides rules of thumb and heuristics, as shown in Table 12-6 and Table 12-7, that allow designers to make a preliminary design for reactive distillation. In particular, the engineer needs to consider maintenance of good liquid mixing and solid-liquid contact having channels available for vapor flow so that reactive components in the vapor can achieve adequate contact with the catalyst. Catalyst containers with higher mass transfer efficiencies reduce costs because of the reduction in required column height.

Table 12-6 Rules of thumb for reactive distillation

(Source: Fair, J. R., Design Aspects of Reactive Distillation, Chem. Eng. P 158, Oct. [1998]).

1. Use pre-reactor when reaction rate at 80% conversion is more than half the initial rate 2. Optimize pressure based on rate, volatilities and azeotropes 3. Locate the reactive zone based on a nonreactive simulation 4. Optimize amount of catalyst based on conversion, yield and capital and operating costs 5. Determine feed location based on volatilities of reactants

6. Recognize influence of azeotropes on reactant ratio

7. Recognize that reacting reflux ratios will be higher than nonreacting reflux ratios. Use a factor of 1.2−1.4 × minimum ratio

Table 12-7 Designer's checklist for reactive distillation (Source: Fair, J. R., Design Aspects of Reactive Distillation, Chem. Eng. P 158, Oct. [1998]).

1. Define chemistry and thermodynamics of reactions

2. Define reaction kinetics

3. Set separation requirements

4. Define phase behavior

5. Choose operating temperature and pressure combination

6. Choose column internals for reactive and nonreactive zones
7. Simulate low conversion conditions to approximate location of reactive zone

8. Select appropriate amount of catalyst

9. Calculate tentative column diameter and height

10. Determine feed stage location

11. Determine stages and reflux relationships

12.11.2. Phase Diagrams and Degrees of Freedom

Consider a single chemical reaction as expressed by:

(12-10)

where A_i represents component i and is the stoichiometric coefficients for component i. By convention, for reactants, for products and if the component i does not take part in the reaction, *i.e.* inert components or solvents. If the reaction is in the liquid phase, this assumption has no effect on the equilibrium condition but will influence the kinetic modeling. The general equations for phase and reaction equilibrium are:

Phase Equilibrium

(12-11)

Reaction Equilibrium

(12-12)

In addition, the mole fractions in the liquid and vapor phases are given by:

(12-13)

A degree of freedom analysis shows that there are 2C + 2 variables related by C + 3 equations (Eqs. (12-11) and (12-14) and 12-14). This leaves C-1 degrees of freedom in agreement with the Gibbs phase rule for reacting systems [262].

(12-14)

where R is the number of independent chemical reactions.

For two phases (P = 2) and one chemical reaction (R = 1), the degrees of freedom are F = C – 1. This indicates the fact that we must specify one less variable than for the corresponding calculation without chemical reaction because we have one additional equation (12-12).

The liquid is treated as a nonideal mixture, using the activity coefficient formalism;

(12-15)

where

= chemical

potential of pure liquid i at the temperature of the mixture. (The pressure dependence is neglected for liquids)

= chemical potential of component i

= mole fraction of component i in the liquid phase

= fugacity coefficient of component i

R = Universal gas constant

T = absolute temperature.

The resulting equilibrium equations are:

Phase Equilibrium

(12-16)

Reaction Equilibrium

(12-17)

where is the number of reactants and is the number of products.

If represents the number of inert components (i.e. these components do not take part in the chemical transformation but they do affect the phase equilibrium and therefore the reaction equilibrium), then

.

Reaction Equilibrium constant,

The reaction equilibrium constant is defined by:

(12-18)

where

The standard Gibbs free energy of reaction represents the difference between the free energy of the products and the free energy of reactants at temperature T. Since and typically have the same sign and the same order of magnitude the following equations apply:

Exothermic Reaction

(12-19)

Endothermic Reaction

(12-20)

Table 12-8 shows typical values for K_{eq} for a selection of reactions.

Table 12-8 Selected reactions and equilibrium constants (Source: Doherty, M. F. and M. F. Malone, Conceptual Design of Distillation Systems, McGraw-Hill, New York, 2001 [262]).

Reaction	Temperature (°C)	Equilibrium constant (K_{eq})	References
Methanol + acetic acid ⇌ methyl acetate + water	$30 < T < 120$	$30 > K_{eq} > 17$	Song et. al. (1998)
Isopropanol + acetic acid ⇌ isopropyl acetate + water	$50 < T < 110$	$K_{eq} \sim 8.7$	Lee and Kuo, (1996)
Butanol + acetic acid ⇌ butyl acetate + water	$100 < T < 120$	$15 > K_{eq} > 11$	Venimadhavan et. al (1999)
2 Methanol ⇌ dimethyl ether + water	$62 < T < 100$	$87 > K_{eq} > 58$	Nisoli et. al. (1997)
Methanol + isobutene ⇌ MTBE	$-10 < T < 67$	$1911 > K_{eq} > 53$	Colombo et. al. (1983)
Methanol + 2 methyl-1-butene ⇌ TAME	$30 < T < 100$	$102 > K_{eq} > 9$	Syed et. al. (2000)
Methanol + 2-methyl-2-butene ⇌ TAME	$30 < T < 100$	$8 > K_{eq} > 1.3$	Syed et. al. (2000)
Methylene glycol ⇌ formaldehyde + water	$T \sim 20 – 25$	$K_{eq} \sim 5 \times 10^{-4}$	Xu et. al. (1985)
Ethylene oxide + water ⇌ ethylene glycol	$T \sim 50$	$K_{eq} > 1000$	
Acetic anhydride + water ⇌ 2 acetic acid	$T \sim 50$	$K_{eq} > 1000$	

12.11.3. A Case Study

Manufacture of Methyl Acetate

Methyl acetate is an intermediate product in the manufacture of a variety of plastics and polyesters (e.g. cellulose acetate, photographic film, Estron acetate for filter tow and textile yarns). The process

involves the esterification of acetic acid [illegible] with methanol [illegible], which is difficult because of reaction equilibrium limitations, the difficulty of separating acetic acid and water, and the presence of azeotropes, both between methyl acetate and methanol, and between methyl acetate and water.

[illegible]

The thermodynamic equilibrium constant [illegible] of this reaction is of the order of 20, and the reactor effluent contains a large amount of all four components. Conventional processes use one or more liquid phase reactors with a large excess of one of the reactants [illegible] in order to achieve high conversion of the other reactant [illegible]. A conventional separating system is described by Siirola [342], which requires eight distillation columns, one liquid extractor and one decanter to break the azeotropes. This process incurs high energy costs with a large capital investment and a large inventory of solvents.

Agreda and Partin [343] invented an alternative process that involves both reaction and separation taking place in the same vessel. High purity methyl acetate is produced with no additional purification steps, and with no unconverted reactant streams to be recovered. This process provides a method of driving the equilibrium limited reaction to high conversions without a large excess of one of the reactant being necessary. This is achieved by allowing the liquid-phase reaction mixture to flash off methyl acetate thereby increasing the conversion.

The feeds to the column are balanced stoichiometrically (e.g. have a molar ratio of 1:1) The light reactant methanol is fed to the bottom section and the heavier acetic acid is fed to the top, so providing good contact between the reactants as a counter current is setup.

The use of excess methanol (e.g. a molar ratio of about 1.1: 2.0) is referred to as “approximate stoichiometric” quantities and the excess methanol is removed from the bottom of the column with the waste water, which is recycled to the feed following purification.

The acetic acid is used both as a reactant, and as an extractive agent for breaking the azeotropes between methyl acetate and methanol, and between methyl acetate and water.

The residence time of the reaction mixture in the column is defined as the clear liquid hold-up volume in the reaction trays divided by the sum of the volumetric flow rates of acetic acid and methanol. The minimum required residence time upon the catalyst concentration, and the number of stages. High reactant conversions and high purity product cannot be achieved with times any less than this minimum. Agreda and Partin obtained a residence time of about 2.4 hours for a catalyst feed rate of about 1 kg of sulfuric acid per 100 kg of acetic acid. This residence time is enough to obtain a high purity methyl acetate with high conversion (∼99 wt %) of the reactants. In this process, a control scheme is provided that allows the attainment of high distillate purity and low reactant losses.

The reactants flow in a countercurrent through the reactor/column, simultaneously reacting and flashing at each stage. The removal of methyl acetate in preference to the other components increases the extent of reaction achieved in each stage. Operating conditions to provide intimate contact in the column between are chosen acetic acid and methanol, between acetic acid and the methyl acetate/water

azeotrope, and sulfuric acid and the methyl acetate/methanol azeotrope. The provision of intimate contact between acetic acid and each of the components in the column results in both high conversion of the reactants and high yield of methyl acetate. This is continuously removed from the top of the column and water is continuously removed from the bottom. Figure 12-17 shows a photograph of an Eastman Company reactive distillation column for methyl acetate production.

Figure 12-17

A photograph of a reactive distillation column for the production of methyl acetate

(by permission of Eastman Co., All rights reserved).

Figure 12-17a shows a schematic of the reactive distillation column. There are four zones in the column, which ensure high conversion. They are a methanol/water stripping section, a reactive distillation section, an extractive distillation section, and a methyl acetate/acetic acid rectification section. Acetic acid is separated from methyl acetate at the top of the column (zone I). In zone II, the acetic acid feed extracts water from methyl acetate. The reaction takes place in the middle of the column (zone III) below the sulfuric acid catalysts feed point as shown in Figure 12-17a. As methyl acetate is formed, it produces a minimum boiling azeotrope with methanol (Figure 12-18). The azeotrope is the lightest boiler in the mixture and is produced from the top of the reactive zone. At the bottom (zone IV), methanol is fed in and also stripped from the descending byproduct water.

Figure 12-17a Reactive distillation column for production of methyl acetate.

Figure 12-18

VLE diagram for the mixture methyl acetate and methanol at 1 atm.

The lower section of the column below the methanol feed performs a methanol-water separation to give high purity water as bottoms product. The top section of the column above the sulfuric acid feed point performs an extractive distillation using the acetic acid as an entrainer to “break” the methyl acetate-methanol azeotrope, and high purity methyl acetate is produced as the distillate.

The proper design of the column internals is essential in order to achieve a high hold-up time. This involves the proper design of the trays, weirs, bubble caps *etc*. The high hold-up times are achieved through the use of reverse flow trays having high weirs, high bubble cap risers and large inlet and flow reserving zone sumps. Composition and temperature profiles for the column as reported by Agreda *et al.* [344] are shown in Figure 12-19.

Figure 12-19

Reactive extracting distillation for methyl acetate production; (a) composition profile; (b) temperature profile
(source: V.H. Agreal *et al.* Chem. Eng. Prog. No. 2, 40–46, [1990]).

The column is an entire chemical plant in one unit producing 400 million pounds per year (280 kmoles/h) of high purity methyl acetate. The process is highly economical and is reported to need only one-fifth of the capital investment of the conventional process and also consumes one-fifth of its energy.

Chemical Reaction and Equilibrium Conversion

Reactive distillation involving methanol , acetic acid to give methyl acetate and water .

The chemical reaction involving these components is:

(12-22)

The pure components and azeotropic boiling data for the mixture at 1 atmosphere are:

Pure component	**Temperature °C**
Methanol	64.5
Acetic acid	118.5
Methyl acetate	57
Water	100.0

Azeotropes

Methyl acetate + Methanol	Minimum boiling azeotrope at 53.9°C
Methyl acetate + Water	Minimum boiling azeotrope at 56.1°C

A pseudo-homogeneous rate model is:

(12-23)

or

(12-24)

where the reaction equilibrium constant K_{eq} as given by Song *et al.* [345] is:

(12-25)

and k_f, the forward rate constant is:

(12-26)

The normal boiling point of methyl acetate for the calculation of k_o from is given a value of 5.1937 hr^{-1}. The heat of reaction is

, which is slightly exothermic. Production is 400 × 10^6 lb/year (280 kmol/h) of methylacetate (Agreda *et al.* [343]).

Through it continuous removal of reaction products, reactive distillation has been successfully applied in improving the economic operation of processes such as methanol and acetic acid reaction. Methyl acetate is an important oil resin used in the manufacture of artificial leathers.

A Case Study: Reactive Distillation of Methyl Acetate

A reactive distillation column having 15 stages with a total condenser and reboiler is to be designed for the production of methyl acetate, with mass flow rate of 45 kgmole/h. A reflux ratio of 5 mol to 1 of the product is to be used. The feed enters at a temperature of 75°C and a pressure of 101.3 kPa at the 10th stage. The overhead product rate coming out of the column is 20 kg mole/h. The pressures at the condenser and reboiler are 90 kPa and 97 kPa respectively. Simulate:

1. the reactive distillation column, assuming no pressure drop across the reboiler and condenser. Determine the mole fractions of components in both the distillate and the bottoms, and also the compositions, temperature and pressure profiles throughout the column.
2. using the reaction mechanism, the stoichiometric coefficients and the sparse continuation solver in Honeywell UniSim® simulation software, determine the mole fractions in the distillate and bottoms, when the chemical reaction occurs on stages 5–10 of the column.

The mole fractions of the components in the feed are:

Methanol	0.4
Acetic acid	0.4

Methyl acetate	0.1
Water	0.1

Operating conditions are:

Feed stream conditions

Stream Name	**Feed**
Temperature	75°C (165°F)
Pressure	101.3 kPa (14.7 psia)
Mass flow	45 kgmole/h (100 lbmole/hr)

Composition	**Mole fraction**
Methanol	0.4
Acetic acid	0.4
Methyl acetate	0.1
Water	0.1

Add a **Distillation Column** with the following connections:

Column name	**Reactive distillation**
Number of stages	15
Feed	Feed, stage 10
Condenser Type	Total
Overhead liquid	Distillate
Bottoms liquid	Bottoms
Condenser Energy	Cond Q
Reboiler Energy	Reb Q
Pressure	

Pressure	
Delta P, Condenser and Reboiler	0 kPa (0 psi)
Condenser	90 kPa (13 psia)
Reboiler	97 kPa (14 psia)
Specification	
Reflux ratio	5
Distillate rate	20 kg mole/hr (44 lb mole/h)

Solution

Honeywell UniSim® Simulation software (reactive-distillation.usc) is used to simulate the reactive distillation of methyl acetate from methanol and acetic acid. Figure 12-19aFigure 12-19b and Figure 12-19c show the compositions, temperature and pressure profiles against tray position from the top. Figure 12-19d shows the process flow diagram with the molar flows and mole fractions of the components in the feed, distillate and bottoms products. The simulation results of the Case Study are shown in Appendix L.

Figure 12-19a Composition vs. tray position from top

Figure 12-19b Temperature vs. tray position from top

Figure 12-19c Pressure vs. tray position from top

Figure 12-19d Process flow diagram of methyl acetate production.

12.12. Advantages and Disadvantages of Reactive Distillation

Advantages

1. In conventional distillation, separate equipment is required for reaction and distillation, but reactive distillation eliminates the use of several piece of major process equipment.
2. Countercurrent flow of reactants results in greater reaction rate than cocurrent.
3. The heat of reaction reduces either condensation duty or reboiling duty, thereby reducing the size of the required equipment.
4. In-situ removal of product can shift an equilibrium reaction to completion.
5. In reactive distillation, less unconverted reactants need to be separated from the products. Thus, reactive distillation requires fewer of stages for the same reflux or less reflux, compared to a conventional distillation. Less amount of reflux decreases the sizes of the condenser and reboiler that are needed.
6. For equilibrium limited reactions, continuous removal of one of the products increases conversion as it converts a reversible reaction into an irreversible one. Thus reactive distillation provides high conversion compared with a conventional reaction.
7. For an exothermic or endothermic reactions, separate cooling

or heating systems are required. In reactive distillation, however, the heat of reaction that is generated can actually be used, either for boiling or for condensation.

8. This reduces operating cost, since the heat of reaction reduces the require load on either the cooling system or the heating system.

9. The process is also made more economical by the lack of need to recycle reactants, and the lower amount of distillate used.

10. For the same reflux ratio, less reflux is required, so both condenser and reboiler heat duties and the power required for pumping the reflux are lower.

11. Efficient heat integration can be applied in the case of exothermic reaction.

Disadvantages

1. In the majority of processes that involve reaction accompanied by distillation, optimum conditions for the reaction and distillation processes are different. To carry out both processes simultaneously in column, a compromise must be made in either reaction or distillation. In the majority of the cases, the compromise is in distillation. Only in a few cases such as methyl tetra butyl ether (MTBE) production are the optimum conditions the same for both reaction and distillation.

2. In conventional distillation, the first step is the reaction, where reactants are brought into contact with solid or liquid catalyst at specified temperature and pressure. This is followed by a separation step, where the desired product is separated from unconverted reactants, catalyst, byproducts and /inerts. This is carried out by distillation. But with reactive distillation with using a solid catalyst, the flowsheet is changed. In the middle section of catalytic distillation column, reaction and distillation take place simultaneously while the upper and lower sections

represent enriching and stripping sections of distillation or they are utilized only for separation.

3. When a reactant is gaseous or very light, the driving force for reaction is lower than a homogeneous liquid-phase reaction at higher pressure, so slowing the reaction down.

4. In these cases the reaction rate may not be compatible with permissible liquid residence time and volumetric holdup for optimal hydraulic operation.

These limitations can be counteracted by proper catalyst design. More active catalysts are required than stand alone reactors.

12.13. Different Ways of Applying Reactive Distillation

Thakore and Bhatt [313] provide different techniques by which reactive distillation can be applied:

Reactive distillation alone: throughout the whole column, reaction and distillation take place simultaneously. Purified product is obtained at one end. This is possible for systems when the pure product point lies within the reaction place.

Reactive distillation plus nonreactive distillation: There are instances for which the pure product point does not lie within the reaction space. For such cases, a nonreactive distillation section (either rectifying, stripping or both) is also required to achieve pure reactants and product.

Reaction alone plus Reactive Distillation plus Nonreactive Distillation (Optional)

This method is applicable if the reaction space required is considerably greater than distillation space.

Sometimes it is not recommended to shift the entire reaction to a

reactive distillation unit because of the following:

1. If the optimum reaction conditions and optimum distillation conditions are different, then it is better to improve only equilibrium conversion in a reactive distillation step, while the major part of the reaction (up to equilibrium conversion) is carried out separately in another reactor.
2. If the existing plant has a number of reactors in series, then only the last, or last few reactors can be changed to reactive distillation unit.

Figure 12-20 shows a comparison between conventional and reactive distillation process.

Figure 12-20

Comparison between conventional and reactive distillation technology.

12.14. Contact Devices Used for Catalytic Distillation

Contact devices are essential to improve the mass transfer rate and to enhance catalyst contact. Various types of contact devices are used for catalytic distillation:

1. Packed tower using random packing elements that contain the catalyst or which are made of the catalyst.
2. A tray column in which the bed of catalyst is placed either in downcomers or above the trays.
3. A packed tower using ordered packing elements which contain catalyst particles. These are sandwiched between two supporting materials in orderly packing elements.

12.14.1. Structured packing

This type of packing has been developed as *KataMax* by Koch Engineering in the U.S.A. and as *KataPak* by Gebrüder Sulzer (Fig. 12.21). The catalyst section in the catalytic distillation column has a conventional corrugated structure with containing wire-cloth distillation packing the appropriate catalyst pellets for the desired reaction. The solid catalyst is held in an envelope of meshed and crimped screen, sealed at the edges. These envelopes are stacked together in modular blocks, so that the crimp of one envelope has a 90° angle of inclination to the crimp of the next. In this fashion, channels are formed for the rising vapor flow and descending liquid flow. At the intersection, they provide intensive mixing of vapor and liquid phases and radial distribution in a block. Gaps are provided between the two layers of catalyst surrounded by screen, referred to as flow channels, which enhance the flow of vapor and where the vapor makes contact with the liquid film.

Figure 12-21

Katapak™ –Structured packing. Used by permission of Sulzer Co.

The catalyst content of *KataMax* or *KataPak* lies between 20 and 25 vol%. For *KataMax* a HETP value (height equivalent to a theoretical plate) between 0.4 and 0.6 m can be used for simulation. For *KataPak*, the HETP is 1.0 m at normal F factors. The specific surface area of *KataMax* is 210 m^2/m^3 and of *KataPak*, 85 to 125 m^2/m^3 (KataPak-S 170.Y or KataPak-S 250 Y). These catalyst types provide good

distribution of liquid and vapor phases at a low pressure drop, efficient contact of the reactants with catalyst pellets, and instantaneous removal of reactants during the distillation process. Thus, it combines excellent separation performance with an efficient mass and heat transfer for chemical reaction.

12.14.2. Bale packing

Here, catalyst particles are enclosed within several pockets in a fiberglass cloth belt. The pockets are normally 25 to 50 mm wide with 6.25 to 12.5 mm web between them. Open ends of pockets are sewn closed. Belts are rolled up between alternating layers of stainless steel mesh. The mesh increases the structural strength of the cloth belt.

Bales constructed for commercial use are roughly cylindrical in shape, and of 200 to 350 mm diameter and 500 mm in height. Several bales are packed tightly into a column such that the pockets remain vertical. Flow channels enhance the contact between vapor and liquid flows, although contact between liquid reactants and catalyst takes place in pockets while vapor-liquid contact occurs in the flow channel. The height equivalent to theoretical plate (HETP) as well as the pressure drop relations for bale packings are supplied by the manufacturer. Figure 12-22 shows the structured bale packing. This type of spatial configuration in a distillation column was first developed and commercialized by CR & L (Chemical Research & Licensing Company) in 1980, however it is now licensed by CD Tech (Chemical Distillation Technologies), a joint venture between CR & L and ABB Lumbus Crest. In this technology, the catalyst is supported in a fiberglass cloth, which is wrapped with stainless steel wire mesh. The demister wire mesh has two tasks. Firstly, it stabilizes the fiberglass packing, and secondly, it provides the void space necessary for distillation. These fiberglass cloths are rolled into catalyst bales and stacked on sieve trays as shown in Figure 12-23.

Figure 12-22

Bale packing

(source: Thakore, S.B., and B.I. Bhatt, Introduction to Process Engineering and Design, McGraw-Hill Publishing Co. Ltd., 2007).

Figure 12-23

Catalytic distillation Tech catalyst bales

(source: M. Sakuth, et al., Reactive Distillation, Ullmann's Processes and Process Engineering, Wiley-VCH, 2004).

12.15. Flowchart for Process Development

Figure 12-24 shows a flowchart that is a conceptual basis for the development of a catalytic distillation process. The essential feature of the chart is the feasibility study, which is followed by pilot plant experiments (optional) and a basic engineering concept. After employing the above part of the chart for process synthesis and design, the parameters of a kinetic model must be determined in an appropriate experimental setup.

Figure 12-24

Flowchart for the process development of a catalytic distillation technology (CSTR = continuous stirred tank reactor).

(source: Michael Sakuth, *et al.* (Reactive Distillation), Ullmann's Processes and Process Engineering, Wiley-VCH, 2004).

Simulation studies with a simple equilibrium stage model (RADFRAC) should provide the position of the reaction zone, the operating conditions in the column, and the optimal feed point to the catalytic distillation column. For an initial investigation, these studies can be carried out by using equilibrium constants to incorporate the chemical reaction into the column. For a detailed investigation, an appropriate reaction rate expression with the determined kinetic parameters should be used.

In pilot plant experiments carried out in a catalytic distillation column with an inner diameter of 50–100 mm, the simulation studies are tested, and as the process is iterative, the results are required to verify and to improve the simulations. The final step of the feasibility study involves feeding the results into a final simulation to decide whether a reactive distillation or conventional setup (e.g. reaction section followed by separation steps with internal recycle streams) should be applied. Preferably a rate-based model (RATEFRAC) should be used to check whether any mass-transfer limitations interfere with the removal of reactants, as such limitations can cause undesired consecutive reactions in the reaction zone [346].

Nomenclature

K_{eq} = Reaction equilibrium constant

k_f = Forward rate constant

k_r = Reverse rate constant

R = Universal gas constant

T = Absolute temperature

= Mole fraction of component i in the liquid phase

Y_i.. = Mole fraction of component i in the vapor phase

= The standard Gibbs free energy of reaction

= Standard enthalpy of reaction

= Chemical potential of pure liquid i at the temperature of the mixture. (The pressure dependence is neglected for liquids)

= Chemical potential of component i

= Fugacity coefficient of component i

Chapter 13. Part 3: Mechanical Designs for Tray Performance

Determining the number of theoretical and actual trays in a distillation column is only part of the design necessary to ensure optimum system performance. The interpretation of distillation, absorption, or stripping requirements into a mechanical vessel with internal components (trays or packing, see Chapter 14) requires use of theoretical and empirical data. The costs of this equipment are markedly influenced by the column diameter and the intricacies of the trays, such as caps, risers, weirs, downcomers, perforations, *etc.* Calculated tray efficiencies in actual can be lost operation by any unbalanced and improperly designed tray.

Contacting Trays

The particular tray selection and its design can materially affect the performance of a given distillation, absorption, or stripping system. Each tray should be designed to give as efficient a contact between the vapor and liquid as possible, within reasonable economic limits. It is impractical in most cases to change the design of each tray to fit calculated conditions. Therefore, the same tray design is usually used throughout the column, or the top section may be of one design (or type) while the lower section is of another design. The more individual tray designs included in a column, the greater the cost.

This concept has not gained commercial popularity since the proprietary nature of the Fractionation Research, Inc. (FRI) data means that it is limited to member organizations, and open literature does not contain much independent research and application data. General industrial and commercial proprietary designs are listed in Table 13-1, but this may not be all-inclusive.

Table 13-1 General Listing and Comparison of the Major Contacting Trays[+]

Compiled from recent manufacturer's literature. Credit is acknowledged for the us of this material, which is not all-inclusive in this table.

Type tray	Manufacturer	Efficiency capacity	Turndown	Pressure drop	Operating flexibility
Bubblecap Figure 13-1Figure 13-2 and Figure 13-4	Several	Med. to 60%	Good low flows, exceeds valve tray	High	Good
Sieve Figures 13-5A, 5C, 6B	Several	High	High/limited	Medium	Medium
Sieve	Linde	High	High/medium	Low	Good/mediu
Dual flow[*] Figure 13-6A	Shell Dev.	High	High/medium	Low	Low
Float valve Figure 13-7A and Figure 13-7B	Nutter	High	High/high	Medium	Good
Fixed V-grid Figure 13-8	Nutter	Outperforms sieve trays	High/high	Medium	Good
MVG tray Figure 13-9A and Figure 13-9B	Nutter	High	High/4:1 > by 15-/ratio 30%	Medium	Good
Flexitray Type A Figure 13-10 valve lift w/srp edge orifice	Koch	Higher than sieve tray, Lower entrain than sieve	Wide range	Low	Good

orifice, round		sieve			
Flexitray Type A w/contoured hole Figure 13-11B	Koch	Higher than sieve tray, Lower entrain than sieve	Wide range		Good
Flexitray Type T, rnd. Figure 13-11A retained by fixed hold down, srp edge hole	Koch	Higher than sieve tray, Lower entrain than sieve	Wide range		Good
Flexitray Type T_o lift, contoured hole Figure 13-10	Koch	Higher than Sieve, w/lower entr.	Wide range	Low	Good
Flexitray Type S, fixed triangular valve Figure 13-10	Koch	Medium	Good, fixed by design	Medium	Good to medium
Ballast, valve V-series V-1 thru V-5 valves Figures 13-12 and 13-15	Glitsch	Higher than sieve	High/high	Medium similar to sieve	Good
Ballast, valve Figure 13-13 and					

13-13 and Figure 13-14 A - series lift, A-1 A-5w/cages Figure 13-13	Glitsch	Higher than Bubble cap or sieve	High/high	Medium to low	Good
Nye tray Figure 13-16**	Glitsch	Same or better than sieve/valve	Similar to sieve or valve	Low	Good
Bubble cap (FRI plain 3, in. and 4 in.) Figure 13-17	Norton	Good	High/high	Medium	Good
Valves, MR2L Figure 13-18	Norton	Good	Medium/good	Medium	Good
Valve, L	Norton	Good	Medium/good	Medium	Good
Caged, MR2	Norton	Good	Medium/good	Medium	Good
Valve, M	Norton	Good	Medium/good	Medium	Good
Cage, MR7	Norton	Good	Medium/good	Medium	Good
Caged, G	Norton	Good	Medium/good	Medium	Good

[#]This specialized Sieve Tray design is of high efficiency and operates with exceptional short tray spacings, sometimes as low as 6 in. between trays.

*Not a wide usage.

**Nye Tray, 10–20% increased tray (over sieve or valve) capacity and good efficiency. More capacity from existing column. Improved inlet area for sieve or valve tray with greater area for vapor–liquid disengagement.

[+]Not offered as all inclusive analysis, but as somewhat general guidelines based on the respective manufacturer's literature description and best interpretation by E. E. Ludwig. This table is not intended to be a decision-making guide, and E. E.

Ludwig recommends that the engineer discusses and presents separation requirements to the respective manufacturers. There is no intension to suggest a negative performance by any manufacturer's design of fabricated equipment.

13.1. Tray Types and Distinguishing Application Features

Bubble Cap

Vapor rises up through “risers” or “up-takes” into a bubble cap, then out through slots as bubbles into the surrounding liquid on the tray. Bubbling action effects contact. Liquid flows over caps, outlet weir and downcomer to the tray below, as shown in Figure 13-1Figure 13-2Figure 13-3Figure 13-4Figure 13-5AFigure 13-5BFigure 13-5C and Figure 13-5D, 13-17 and 13-19.

Figure 13-1

Bubble cap tray schematic – dynamic operation.

Figure 13-2

Bubble cap tray in large column.

Used by permission, Glitsch, Inc.

Figure 13-3

Slip-type cartridge assembly for bubble cap trays in small column, 1-ft-10 -in. I.D.

Used by permission, Glitch, Inc.

Figure 13-4

Bubble cap performance.

Figure 13-5A

Sieve or perforated tray with downcomers.

Figure 13-5B

For sieve tray layout arrangements; typical one-, two-, and four pass tray flow patterns with clarifying flow markings by Ludwig.

Used by permission, Glistch, Inc., Bul. 4900 – 5th Ed.

Figure 13-5C

Linde multiple-downcomer tray.

Used by permission, Union Carbide Corp., Linde Division.

Figure 13-5D

Ripple tray (no downcomers).

Used by permission, Stone and Webster Engineering Corp., Boston, Mass.

Capacity: moderately high, maintains efficiency.

Efficiency: as high as other tray designs.

Entrainment: about three times that of perforated type plate or sieve tray. Jet-action accompanies bubbling.

Flexibility: the most flexible of tray designs for high and low vapor and liquid rates. Allows positive drain of liquid from tray. Liquid heads maintained by weirs.

Application: all services except extreme coking, polymer formation or other high fouling conditions. Used for extremely low flow conditions where tray must remain wet and maintain a vapor seal.

Tray Spacing: 18-in. average, 24-to 36-in. for vacuum conditions.

Sieve Tray or Perforated Tray With Downcomers

Vapor rises through small holes (⅛ to 1-in.) in tray the floor, bubbles through the liquid in a fairly uniform manner. Liquid then flows across tray floor over weir (if used), through downcomer to the tray below (Figure 13-5AFigure 13-5B and Figure 13-5C and 13-6B).

Capacity: As high or higher than bubble cap at design or down to 60% of design rates with good efficiency. At lower throughputs, performance drops since efficiency falls off rapidly.

Efficiency: As high as bubble caps in the region of design, but falls to unacceptable values when capacity reduces below 60% (approximately).

Entrainment: Only about one-third that of bubble cap trays.

Flexibility: Not generally suitable for columns operating under variable load, falling below 60% of design. Tray weeps liquid at low vapor rates.

Application: Systems where high capacity, near-design rates to be maintained in continuous service. Handles suspended solid particles by flushing them down from tray to tray. Holes become plugged in salting-out systems where trays run hot and dry (as underside of bottom tray).

Tray Spacing: Can be closer than bubble cap due to improved

entrainment. Fifteen inches is average, 9-in., 10-in. and 12-in. are acceptable, with 20-to 30-in. for vacuum.

Perforated Plate Without Downcomers: (Dual-Flow, from FRI)

Vapor rises through holes (- to 1-in.) in the tray floor and bubbles through liquid. At the same time the liquid head forces liquid countercurrent through these holes and onto tray below. Liquid flow forms random patterns in draining and does not form continuous streamlets from each hole. See Figures 13-5D and 13-6A.

Figure 13-6A

Perforated tray without downcomer.

Used by permission, Hendrick Mfg. Co., Carbondale, Pa.

Figure 13-6B

Sieve tray with integral downcomer.

Used by permission, Hendrick Mfg. Co., Carbondale, Pa.

Figure 13-7A

Nutter BDH™ Float valve tray with downcomer.

Used by permission, Nutter Engineering, Harsco Corp.

Figure 13-7B

Nutter Float valves™ in open and closed positions for use on distillation trays.

Used by permission, Nutter Engineering, Harsco Corp. Bul. N-2.

Figure 13-8

Nutter V-Grid™ fixed valve for trays.

Used by permission, Nutter Engineering, Harsco Corp. Bul. N-2.

Figure 13-9A

Nutter MVG™ high-performance fixed valve with 4:1 turndown ratio. Used in new installations and to replace sieve trays.

Used by permission, Nutter Engineering, Harsco Corp. Bul. CN-4.

Figure 13-9B

Nutter LVG™ long, SVG™ short, and MVG™ tray slots. MVG™ tray slots are always placed in a triangular pattern.

Used by permission, Nutter Engineering, Harsco Corp.

Figure 13-10

Types of standard Koch valves. Used by permission, Koch Engineering, Co., Inc. Bul. KT-6A.

A. Standard valve with integral legs used for most services, utilizing a sharp-edged orifice in the tray floor.

A_o Same "A" valve, but with a contoured hole in the tray floor lower operating pressure drop.

T Round valve retained by a fixed hold down and a sharp edge hole in the tray floor for all services, including fouling, slurry and corrosive applications.

T_o The same "T" valve and hold down with a contoured, low pressure drop hole in the tray floor.

S Stationary valve punched up from the tray floor.

The A, A_o, T and T_o valves can be supplied either with a flat periphery for tightest shutoff against liquid weepage at turndown rates or with a three-dimpled periphery to minimize contact with the tray deck for fouling or corrosive conditions.

Figure 13-11A Type “T” Flexitray®.

Used by permission, Koch Engineering Co. Inc.

Figure 13-11B

Type “A” Flexitray® with double pass.

Used by permission, Koch Engineering Co. Inc.

Figure 13-12

Glitsch Ballast® Valves, V-series. Used by permission, Glitch, Inc., Bul. Bu-69 (Rev.). (A) Standard Dimensions: Diameter: 1⅞“ Initial Opening: 0.10” V-1, V-4, GV-1 and H-1H. (B) Standard Dimensions: Diameter: 1⅞“ Initial Opening: none (flush seating), V-IX, V-4X, GV-1X and V-1HX. (C) V-1 Type (Flat orifice) Diameter of Orifice Opening: 1 ”. (D) V-4 Type (Extruded Orifice).

Figure 13-13

Glitsch Ballast® Valves, Aseries.

Used by permission, Glitch, Inc., But. Bu-69 (Rev.).

Figure 13-14

Type A-1 Ballast® Tray.

Used by permission, Glitch, Inc.

Figure 13-15

Type V-1 Ballast® Tray.

Used by permission, Glitch, Inc.

Capacity: Quite similar to sieve tray, as high or higher than bubble cap tray from 50% up to 100% design rate (varies with system and

design criteria). Performance at specification quality falls off at lower rates.

Efficiency: Usually not quite as high as bubble caps in the region of design, but falls to unacceptable values below 60% of the design rate.

Entrainment: Only about one-third that of bubble cap tray.

Application: Systems where high capacity near-design rates to be maintained in continuous service. Handles suspended crystal and small solid materials, as well as polymer forming materials. Holes become plugged in salting-out systems where trays run hot and dry (as underside of bottom tray), Good in vacuum or low-pressure-drop design.

Tray Spacing: Can be closer than bubble cap due to improved entrainment; 12-inch is average; 9 to 18-in, acceptable; 18 to 30 in, for vacuum.

Proprietary Trays

There are many special tray designs which solve specific problems and exceed the capabilities of conventional trays. The comments regarding performance are those claimed by the manufacturer, see Table 13-1.

Not all tray designs solve special problems, even though some may have unique performance features. Most of these trays have the bubbling contact action from valves, either moveable (liftable) or fixed (usually stamped or cut from the tray floor itself). In addition to the valve trays above, there are sieve trays (with multiple downcomers) from Union Carbide Corp., Linde Div., and the Turbogrid tray of Shell Development Co., and Ripple Tray (sieve type) of Stone and Webster Engineering Corp.

13.2. Bubble Cap Tray Design

The bubble cap has been studied extensively, and several design recommendations have been presented over the years. The most complete and generally applicable is that of Bolles [5]. It should be stressed that proper mechanical interpretation of process requirements is essential in the design for efficient and economic operation. There is not just one result, but a multiplicity of results, each unique to a particular set of conditions, and some more economical than others. Yet at the same time, many of the mechanical design and fabrication features can be identical for these various designs.

The tray and caps operate as a unit or system; therefore they must be so considered in the design (Figure 13-1Figure 13-2Figure 13-3 and Figure 13-4).

Due to the public nature of so many design techniques, the engineer has sufficient information to prepare a design that can be expected to perform satisfactorily.

Additionally, nearly all of the major tray manufacturers can and do design bubble cap trays as well as the other types on request.

The Fractionation Research, Inc. design procedures are proprietary for the financial participants in this extensive test program.

Standardization

The custom design of the trays for each application is usually unnecessary and uneconomical. Instead most designers use a standard reference tray layout and cap size to check each system. If the results of the tray hydraulics study indicate unsatisfactory operation for the standard tray, then alterations of those features controlling the out-of-line performance is in order, using the same method as will be outlined for the initial design of a custom tray. It is understood that

such a standard tray cannot be optimum for every application but experience has demonstrated that many applications satisfy the standard tray type. The economic advantages of using a limited number of bubble cap sizes and designs are reflected in warehouse stocks. The standardization of layouts, downcomer areas, weir lengths and many other features are reflected in savings in engineering mechanical design time. At the same time, systems that do not adapt themselves to this standardization should be recognized and handled as special designs.

Design Objectives

Each tray design should ultimately resolve and achieve the following:

1. *Capacity*: high for vapor and/or liquid as required. This yields the smallest column diameter for a given throughput. Flexibility or adaptability to high and low fluctuations in vapor and liquid rates.
2. *Pressure Drop*: low pressure drop is necessary to reduce temperature gradients between the top and bottom of the column. High pressure drop is usually (but not always) associated with uneconomical design. In some systems, pressure drop is not a controlling feature within reasonable limitations.
3. *Efficiency*: high efficiency is the objective of each tray performance. The better the contact over a wide range of capacities, the higher will be the efficiency throughout this range.
4. *Fabrication and Installation Costs*: details should be simple to maintain low costs.
5. *Operating and Maintenance Costs*: mechanical details must account for the peculiarities of the system fluids (coking, suspended particles, immiscible fluids, etc.) and accommodate the requirements for drainage, cleaning (chemical or mechanical), corrosion, etc., in order to keep the daily costs of operation and downtime to a minimum.

13.3. Bubble-Cap-Tray Tower Diameter

Column diameter for a particular service is a function of the physical properties of the vapor and liquid at the tray conditions, efficiency and capacity characteristics of the contacting mechanism (bubble trays, sieve trays, etc.) as represented by velocity effects including entrainment, and the pressure of the operation. Unfortunately the interrelationship of these is not clearly understood. Therefore, diameters are determined by relations correlated by empirical factors. The factors influencing bubble cap and similar devices, sieve tray and perforated plate columns are somewhat different.

The Souders-Brown [67] empirically correlated maximum allowable mass velocity is represented in Figure 13-20 for "C" Factor determination, and in Figure 13-21 M for solution of the relation:

(13-1)

where

W = maximum allowable mass velocity through column using bubble cap trays, lb/(ft^2 crosssection) (hour)

C = factor from Figure 13-20 related to entrainment

ρ_v = vapor density, lb/ft^3

ρ_L = liquid density, lb/ft^3

Figure 13-16

Glitsch Nye™ Tray action to improve conventional sieve and valve tray performance by 10–20%.

Used by permission, Glitch, Inc. Bul. GLI-5138.

Figure 13-17

Norton FRI Plain Bubble Cap (3 in. and 4 in.), slotted skirt caps available.

Used by permission, Norton Chemical Process Products Corp., Stow, Ohio.

Figure 13-18

Typical Norton Valve Tray Valves.

Used by permission, Norton Chemical Process Products Corp., Stow, Ohio., Bul. FT-2.

Figure 13-19

Typical 3⅞-in. I.D. (nominal 4 in.) pressed bell bubble cap.

Figure 13-20

"C" factors for column diameter using bubble cap trays.

Adapted by permission. The Americal Chemical Society, Souders, M., jr., and G. G. Brown, Ind. and Eng. Chem. V., 26, p. 98 (1934), All rights reserved.

Figure 13-21

Allowable mass velocity for fractionation, absorption and stripping columns.

The "C" factor is determined at the top and bottom (or intermediate) positions of the column in order to evaluate the point of maximum required diameter. The "W" rate obtained in this solution is the maximum that is allowable, and hence corresponds to the minimum acceptable diameter for operation with essentially no entrainment carryover from plate to plate. Normally a factor of "safety" or "ignorance" of 1.10 to 1.25 would be applied (i.e. W divided by 1.10 to 1.25) if irregularities in capacity, system pressure or other significant variables can be anticipated. Recent experience indicates that the relation is somewhat conservative for pressure (5 to 250 psig) operated distillation systems, and the maximum allowable rate can be increased by 5-15% (W times 1.05 to 1.15) exercising judgment and caution. Basically this reflects satisfactory operation at conditions tolerating some entrainment with no noticeable loss in fractionating efficiency. In any case, the shell diameter should be rounded to the

nearest inch on the diameter for fabrication standardization. Diameters such as 3 ft, 8⅝ in. inside diameter are to be avoided, but can be used if conditions warrant. Standard tray layouts for caps, weirs, *etc.* are usually set at 6 in. intervals of diameter starting at about 24 to 30 in.

The diameter based on vapor flowrate, V′, in the region of greatest flow:

(13-2)

Entrainment may not be the controlling factor in proper design. In cases of high liquid load or with extremely foamy or froth fluids the tendency to flood is generally increased by close tray spacing. The hydraulics of tray operation must be evaluated and the liquid height in the down comer reviewed to control flooding. If liquid height in the downcomer exceeds one-half the tray spacing, the spacing should be increased and the column rechecked. In such cases entrainment is of no worry, as the allowable entrainment vapor capacity will be greater than is needed to satisfy the increased tray spacing.

13.4. Tray Layouts

Flow Paths

The simplest tray arrangement when fluid flow and mechanical details are taken into consideration, is the cross-flow design shown in Figure 13-22. It fits the majority of designs, although when liquid flows become small with respect to vapor flows the reverse flow tray is recommended, and when liquid load is high with respect to vapor, the double pass tray is suggested, as the path is cut in half and the liquid gradient reduced. For extremely high liquid loads the double-pass cascade is suggested. These last two configurations are usually encountered only in large-diameter towers.

Figure 13-22

Tray types by liquid paths.

The liquid flow paths across the tray are important, as channeling to one area or another prevents efficient vapor contact. Short tray baffles are often installed to prevent short circuiting, particularly near the column shell wall. Segmental downcomers with straight chordal type weirs provide an efficient initial distribution pattern for the liquid. Circular or pipe-type downcomers with correspondingly shaped weirs require careful attention to the liquid path as it leaves or enters such a downcomer. For small liquid flows they serve very well. A guide for provisional selection of the tray type for a given capacity is given in Table 13-2 [5].

Table 13-2 Guide for Tentative Selection of Tray Type

Used by permission, Boles, W. L., *Petroleum Processing*, Feb–May (1956).

Estimated tower dia.ft.	Range of liquid capacity; GPM			
	Reverse flow	Cross flow	Double pass	Cascade Double-pass
3	0–30	30–200	–	–
4	0–40	40–300	–	–
6	0–50	50–400	400–700	–
8	0–50	50–500	500–800	–
10	0–50	50–500	500–900	900–1,400
12	0–50	50–500	500–1,000	1,000–1,600
15	0–50	50–500	500–1,100	1,100–1,800
20	0–50	50–500	500–1,100	1,100–2,000

Figure 13-23 and Table 13-3 identify the distribution of areas of a tray by the action of the tray area.

Figure 13-23

Classification of tray area.

Used by permission, Bolles, W. L., Petroleum Processing, Feb.–May (1956).

Table 13-3 Approximate Distribution of Areas as Percentage of Tower Area (allocated cap area is determine by difference)

Used by permission, Bolles, W. L., *Petroleum Processing*, Feb.–May (1956).

Tower diameter ft.	Downflow area		Liquid distribution area			
	Cross flow	Double pass	Cross flow	Double pass	Cascade double	End wastage
3	10–20	–	10–25	–	–	10–30
4	10–20	–	8–20	–	–	7–22
6	10–20	20–30	5–12	15–20	–	5–18
8	10–20	18–27	4–10	12–16	–	4–15
10	10–20	16–24	3–8	9–13	20–30	3–12
12	10–20	14–21	3–6	8–11	15–25	3–10
15	10–20	12–18	2–5	6–9	12–20	2–8
20	–	10–15	–	5–7	9–15	2–6

A tray design guide is given in Table 13-4 and is as presented by Bolles [5], except with modifications where noted. Figure 13-24 is a 3 ft 0 in. diameter tray, and shows representative details associated with

tray design. A typical 4 in. pressed cap is shown in Figures 13-17 and 13-19.

Table 13-4 Tray Design Guide for Bubble Caps*

Materials of construction	
Type	Light gage metal†
Material	Determined by corrosion conditions
Tray type	
General use	Cross-flow
Low L/V ratio	Reverse-flow
High L/V ratio	Double-pass
Very high L/V or very large towers	Double-pass, cascade
Downcomers and weirs	
Downcomer type	Segmental
Downflow baffle	Vertical
Weirs for normal loads	Straight
Weirs for low loads	Notched
Weir adjustment	1–3 in. †
Length: cross-flow trays, % tower dia.	60–70%†
Length: double pass trays, % tower dia.	50–60%†
Downcomer width for double-pass trays	8–12 in. †
Bubblecaps	
Nominal size for:	
2.5–3 ft. towers	4 in. †
4–10 ft. towers	4 in.
10–20 ft. towers	6 in.
Design	Use suggested standard†

Used by permission, Bolles, W. L., *Petroleum Processing*, Feb.–May (1956).

Design	Use suggested standard[‡]
Pitch	Equil. Triangular, rows normal to flow
Spacing	1–3 in.
Skirt height	0.25–1.5 in. [‡]
Fastening	Removable design
Clearances	
Cap to tower wall	1.5 in. minimum
Cap to weir	3 in. minimum
Cap to downcomer or downflow baffle	3 in. minimum[‡]
Tray dynamics	
Mean slot opening	
Maximum	100% slot height
Minimum	0.5 in.
Mean dynamic slot submergence	
Vacuum operation	0.25–1.5 in. [‡]
Atmospheric	0.50–2.0 in. [‡]
50–100 psig	1.0–3.0 in. [‡]
200–500 psig	1.5–4.0 in. [‡]
Vapor distribution ratio (Δ/hc)	0.5 maximum
Height clear liquid in downcomers	50% downflow height, maximum
Downflow residence time	5 seconds, minimum
Liquid throw over weir	60% down flow width, maximum
Entrainment	
As mol/mol dry vapor	0.10 maximum
Pressure drop	As limited by process
Tray Spacing	
For towers 2.5–10 ft. [‡]	18 in.
For towers 4–20 ft. [‡]	24 in.

For towers 4–20 ft.	24 in.
Miscellaneous design factors	
Inlet weirs	Use as required for liquid distribution‡
Intermediate weirs	
……..	Minimum height > height liquid downstream
Reverse-flow baffles	
……..	Minimum height twice clear liquid
Redistribution baffles	
Location:	
All rows where end space is 1-in. >	cap spacing
Clearance to caps	Same as cap spacing
Height	Twice height clear liquid
Downflow baffle seal	
Weir to baffle < 5ft.	0.5 in.
Weir to baffle 5–10 ft.	1.0 in.
Weir to baffle > 10 ft	1.5 in.
Tray design deflection (structural)	⅛ in.
Drain holes	
Size	⅜–⅝ in.
Area.	4 sq in./100 sq. ft. tray area
Leakage	Max. fall 1.0 in. from top of weir in
	20 min. with drain holes plugged.
Construction Tolerances	
Tray levelness‡	⅛-inch. Max. under 36-inch. dia. ‡
	-inch. Max. 36–60 inch dia. ‡
	¼ inch Max. over 60-inch. dia‡
Weir levelness …….	± -in. ‡
‡ Indicates modification by Fr. Fr. Lockhart	

*indicates modification by E. E. Ludwig.

Figure 13-24

Typical bubble cap tray details: 3-ft 0 in. diameter column, 4-in. caps.

These figures show only one set of many possibilities for a general purpose tray. The generality of the design does mean that it is not optimized for a specific system, however. Table 13-5 gives bubble cap and riser layout data and weir lengths for other sizes of general purpose trays.

Table 13-5 General Purpose Distillation Trays using 3⅞-in. I.D. Pressed Metal Bell Caps on 5¼-in. Centers

Inside diameter	**2 ft-6 in.**	**3 ft-0 in.**	**3 ft-6 in.**	**4 ft-0 in.**	**5 ft-0 in.**	**6 ft-0 in.**	**6 ft-6 in.**	**7 ft-0 in.**	**8 ft-0 in.**
Total area, ft^2	4.91	7.07	9.62	12.56	19.63	28.28	33.18	38.48	50.26
Liquid flow	Cross	Cross	Cross	Cross	Cross	Cross	Cross	Cross	Cross
No. downflow weirs (1)	One	One	One	One	One	One	One	One	One
No. downflow seals (2)	One	One	One	One	One	One	One	One	One
No. caps & risers/tray	18	27	37	51	79	129	150	173	223
No rows/trays	4	5	6	7	9	11	12	13	15
Total slot area, ft^2 (3)	1.15	1.73	2.38	3.27	5.08	8.29	9.64	11.1	14.29

Percent of tower area	23.4	24.5	24.8	26.0	25.9	29.3	29.1	28.9	28.50
Total riser area, ft^2 (4)	0.683	1.04	1.42	1.96	3.00	4.95	5.56	6.64	8.57
Percent of tower area	13.9	14.7	14.8	15.6	15.3	17.5	16.75	17.25	17.05
Overflow weir length, ft (1)	1.625	2.05	2.35	2.686	3.35	4.0	4.31	4.62	5.21
Percent of tower diameter	65.0	68.0	67.1	67.0	67.0	66.7	66.0	66.05	65.1
Downflow segment area (5)									
Maximum area, ft^2	0.338	0.642	0.80	0.96	1.41	2.12	2.39	2.85	3.50
Minimum area, ft^2	0.167	0.225	0.33	0.45	0.69	0.886	1.027	1.14	1.47
Under flow clearance, in.	2½	2¾	2¾	2¾	2¾	2¾	3	3	3
Under flow area, ft^2	0.274	0.354	0.426	0.473	0.604	0.710	0.750	0.887	1.01
Up flow area,									
Minimum ft^2 (6)	0.137	0.365	0.335	0.552	0.795	0.740	0.882	1.000	1.21
(1) 2½ in. minimum height above tray floor, with adjustable weir 0 to 3½ in. additional.									
(2) 3½ in. high above tray floor.									
(3) Slots are 50⅛ in. × 1½ in. per cap. Caps on triangular pitch.									
(4) Riser inside diameter = 2.68 in.									
(5) Design uses tapered segmental downcomer and inlet weir.									
(6) Area between downcomer and inlet weir for up-flow of inlet liquid. All trays included access manway.									

Caps suitable for particular tray designs are shown in Figure 13-25Figure 13-26 and Figure 13-27. The rectangular caps require layouts that differ from bell caps, but similar design principles for

flow path evaluation.

Figure 13-25

Bubble cap and riser design.

Used by permission, Glitsch, Inc.

Figure 13-26

Bubble cap and riser comparison data.

Used by permission, Glitsch, Inc.

Figure 13-27

Bubble cap and riser – types and hold-downs.

Used by permission, Glitsch, Inc.

13.5. Liquid Distribution: Feed, Side Streams, Reflux

For tray columns, bubble caps, valves or sieves, the feed liquid usually enters the column either in between functioning trays or at the top (reflux). The liquid, or liquid/vapor mixture for flashing liquids, must be dispersed uniformly across the tray. Such an arrangement often requires a special tray, designed to allow the liquid to flow downward to the next active separation tray, and at the same time allow vapor to rise uniformly to the underside (entering) of the next active (working) tray above. This requires attention to the respective flows in order to avoid upsetting the separation performance of the working trays. Allowance should be made for such an upset by adding two actual trays (or minimum of one theoretical tray divided by the estimated tray efficiency) to the actual trays in the column, at an appropriate location.

Feintuch [221] presents calculations for a pipe distributor with tray and downcomers to disperse the reflux or feed liquid uniformly across the tray (which should not be counted as a "working tray", but a distribution device) and allow the liquid to flow from the downcomer to the tray below to be the first "working" tray. The same concept applies to intermediate feed trays. Designs may vary depending on column diameter (see Figure 13-22) and requirements for liquid and vapor flow. Some designs direct the incoming liquid into the downcomer of the top tray of a column and just adjustment the weir heights.

Layout Helps

The scheme for arranging caps and downcomers as proposed by Bolles [5] is one of several that are available. Plain drafting using the suggested guides is a somewhat more involved approach. The layout sheet of Figure 13-28 and Figure 13-29 is convenient once cap-pitch master layouts are made for the sizes normally used. Figures 13-24, Figure 13-29Figure 13-30Figure 13-31Figure 13-32Figure 13-33Figure 13-34Figure 13-35Figure 13-36 and Figure 13-37 present tray layouts matching the data given in Table 13-5. These have

successfully fitted many different processing situations.

Figure 13-28

Distillation tray layout form (use with Figure 13-29, not reproduced to scale).

Used by permission, Wyatt Metal and Boiler Works, Inc © 1956).

Figure 13-29

4-in. diameter bubble caps on 5.5-in Δ pitch.

Used by permission, Wyatt Metal and Boiler Works, Inc © 1956.

Figure 13-30

2-ft-6-in. diameter column – 4 in. caps.

Figure 13-31

3-ft-6-in. diameter column – 4 in. caps.

Figure 13-32

4-ft-0-in. diameter column – 4 in. caps.

Figure 13-33

5-ft-0-in. diameter column – 4 in. caps.

Figure 13-34

6-ft-6-in. diameter column – 4 in. caps.

Figure 13-35

6-ft-6-in. diameter column – 4 in. caps.

Figure 13-36

7-ft-0-in. diameter column – 4 in. caps.

Figure 13-37

8-ft-0-in. diameter column – 4 in. caps.

Cap Layout

Caps should be arranged on the plate in a 60° equilateral triangular layout, with the liquid flowing into the apex of the triangle rather than parallel to the base. The liquid flows are normal to each row of caps.

Cap Pitch Center to Center

Cap sizes 2 in., 3 in., and 4 in.: recommended cap lane distance of 1½ in. (with 1¼-in. minimum, 1¾ inch maximum) plus cap outside diameter. For cap sizes of 6 and 8 in.: recommended cap lane distance

of 1½ in (1½-in. minimum, 2½-in. maximum) plus cap outside diameter.

Weirs

Figure 13-38 is convenient for deriving weir lengths relative to their effect on segmental downcomers.

a. **Inlet**

Figure 13-38

Segmental downcomer design chart.

Used by permission, Bolles, W. L. Pet. Processing; Feb.–May (1956).

These contribute to the uniform distribution of liquid entering the tray from the downcomer. There are about as many tray designs without weirs as with them. The downcomer without inlet weir tends to maintain uniform liquid distribution itself. The tray design having a recessed seal pan ensures against vapor backflow into the downcomer, but this is seldom necessary. It is not recommended for fluids that are dirty or tend to foul surfaces. The same disadvantages apply to the inlet weir. The first row of caps next to the weir or inlet downcomer must be set back far enough to prevent bubbling into the downcomer. The inlet weir prevents this, although it can also be handled by leaving a gap of approx. 3 in. between inlet downcomer and the nearest face of the first row of caps.

The height of an inlet weir, if used, should be 1 to 1½ in. above the top of the slots of the bubble caps when installed on the tray.

If inlet weirs are used they should have at least two slots ¾-in. by 1-in. flush with the tray floor to aid in flushing out any trapped sediment or other material. There should also be weep or drain holes below the downcomer to drain the weir seal area. Their size should be set by the type of service, but a minimum of ⅜-in. is recommended.

b. **Outlet**

Necessary to maintain a seal on the tray, thus ensuring bubbling of vapors through liquid. The deeper the submergence, i.e., the distance between top of slots of bubble caps and the surface of the liquid flowing on the tray, the lower the tray pressure drops. However, this submergence must have some reasonable minimum value (¼ to ⅜-in.) to avoid excessive bypassing of vapor through void spots in the surging, moving liquid body as it travels across the tray from inlet to outlet.

The adjustable weir feature of many tray designs allows a standard tray to be utilized in different services by readjusting the weir height as needed. The fixed portion of the weir should never be lower than the top of the slots of the bubble caps. Depending upon service, the adjustable weir should be capable of being changed by a minimum of 2 in., with designs for some trays being 4 to 6 in.

Downcomer

Figure 13-38 is convenient for determining the downcomer area and width for a given weir length (see Table 13-5).

The downcomer from a tray must be adequate to carry the liquid flow plus entrained foam and froth. This foamy material is disengaged in the downcomer so only clear liquid flows onto the tray below. The vertical and straight segmental downcomer is recommended, although the segmental tapered design has been used quite successfully

(Figures 13-5A and 13-24). In the latter design the wide mouth of the inlet as compared to the outlet is considered to provide better foam disengagement conditions.

The consensus seems to be that a ratio of upper area to the outlet area (lower) needs to be in the range of l.5 to 2.0 [190]. Circular or pipe downcomers are also used, usually for low liquid flow and small diameter columns, generally below 18 in. diameter. The pipe projects above the tray to serve as the overflow weir. (Note: Calculate flow down on this circular basis, assuming the pipe can be surrounded by liquid on the tray).

The size of the downcomer seal on the tray [5] depends on the liquid flow path:

Liquid Path, Downcomer to outlet weir, ft	Downcomer seal, in.
Below 5	0.5
5–10	1.0
Above 10	1.5

13.6. Liquid By-Pass Baffles

Also known as redistribution baffles, these short stub baffles guide the liquid flow to prevent excessive bypassing of the bubble cap field or active tray area. Unfortunately this action is overlooked by many designers with resulting low tray efficiencies. Table 13-4 gives recommendations for layout.

13.7. Liquid Drainage or Weep Holes

Holes for drainage must be adequate to drain the column in a reasonable time, yet not large enough to interfere with tray action. Draining of the column through the trays is necessary before any internal maintenance can be started, or before fluid services can be changed, when mixing is not desirable. The majority of holes are also

placed adjacent to the outlet or downcomer weir of the tray. However, some holes are also placed in the downcomer inlet area, or any suspected low point in the mechanical layout of the column.

The study of Broaddus *et al.* [7] can be used to develop the following drainage time relationship, and is based on fluids of several different densities and viscosities:

(13-3)

where

θ = time to drain tower, minutes

A_n = net open liquid area of one tray, equal to total tower cross section minus area occupied by caps and minus area of segmental or other downcomer at outlet of tray, ft^2

ρ = liquid density, grams/cc at liquid temperature in tower

μ = viscosity of liquid at tower temperature, centipoise

d_h = weep hole diameter, in. Note that this is the diameter equivalent to the area of all the weep holes/tray

= height of overflow weir or bubble cap riser, whichever is smaller, in.

N = total number of actual trays in tower

The accuracy of this relation is given as 14% maximum, 6% average.

A general recommendation [5] is to provide four square inches of weep hole area per 100 ft^2 of net open liquid tray area in the tower. This latter refers to the total of all trays in the tower.

13.8. Bottom Tray Seal Pan

The bottom tray of a tower must have its downcomer sealed to prevent upflow of reboiled vapors. The downcomer of this tray is usually

equal to or 6 in. longer than the other downcomers, to ensure that bottom vapor surges or pulses in pressure do not break the seal. The seal pan is designed to reduce liquid back pressure and minimize restrictions to liquid flow.

13.9. Turndown Ratio

Turndown can be applied to all types/styles of tray columns; however, it is more relevant to sieve and valve trays. The generally accepted explanation of turndown is as follows [199] (also see Figure 13-39).

Figure 13-39

Qualitative effect of liquid and vapor loads on bubble cap tray performance.

Used by permission, Bolles, W. L. Pet. Processing; Feb.–May (1956).

Turndown ratio is the ratio of the maximum allowable vapor rate at or near flooding conditions (rates) to the minimum vapor rate when weeping or liquid leakage becomes significant; it may be termed the minimum allowable vapor velocity [193, 199, 200].

1. For bubble cap trays, this ratio is approximately 10:1.
2. For sieve trays, this ratio is approximately 2-3:1.
3. For valve trays, this ratio is approximately 4-5:1.

13.10. Bubble Caps

Although there are many styles and dimensions of caps in use, the round bell shaped bubble cap is quite practical and efficient. It is recommended as a good basis for fulfilling contacting requirements. This does not infer that other contacting caps are not acceptable; in

fact many are in use in the chemical and petroleum industry. Their design criteria are proprietary and not available to the designer.

Dimensions

The most popular, and perhaps most adaptable size is about 4 in. O.D. (3⅞ in. I.D.). The 3-in. and 6-in sizes are also in common use for smaller and larger diameter towers, although it is not necessary to change cap size with tower diameter. For a given active cap area the cost of installing smaller diameter caps is 10-15% greater than the 4-in., while the larger (and fewer in number) 6-in. caps are about 15% cheaper. Smaller waste tray area is usually needed with the 3-in. caps than the 4-in. or 6-in. caps (Table 13-6).

Table 13-6 General Purpose Guide for Pressed Cap Size Selection

Tower diameter ft.	General purpose size I.D., in.	Possible alternate, in.
2.5, 3	3⅞	2¾; 3; 3⅝
3.5, 4	3⅞	3; 5⅝
5, 6, 7, 8, 9, 10	3⅞	6
11, 12, 13, 14	6	3⅞

For the sake of standardization and developing a feel for the effect of various design variables on tray performance, the 4-in. cap (or 3¾-in. or 4⅜-in.) caps are recommended as good general purpose units. This means that any application from a 2.5-ft dia. to 10-ft dia. tower is first evaluated using the 4-in. cap. If there are points of poor performance, cap sizes can be changed and the performance re-evaluated.

Pressed steel caps of 12 to 14 U.S. Standard gage are the most frequently used, although cast iron caps are used in some services, such as corrosive chlorinated hydrocarbons, drying with sulfuric acid, *etc*. Alloy pressed caps have the light weights desirable for tray

construction, and frequently serve quite well in corrosive conditions. Special caps of porcelain, glass, and plastic are also available for specific applications. Heavier caps require heavier trays, or more supports in the lighter trays. The use of hold-down bars on caps is not recommended for normal installations, instead, individual bolts and nuts are preferred. Some wedge type holding mechanisms are satisfactory as long as they will not vibrate loose.

13.11. Slots

Slots are the working part of the cap, *i.e.* the point where bubbling action is initiated. Slots are usually rectangular or trapezoidal in shape, either one giving good performance. A single study [5] indicates that rectangular slots give slightly greater capacity than trapezoidal, while the latter give slightly better performance at low vapor rates (flexibility). This study shows that triangular slots are limited in capacity, hence the rectangular slots are preferred.

Slot Sizes

Width: ⅛-½-in., ⅛-in. recommended rectangular ⅛-in. x ¼-in. to ¼-in. x ¾-in., in. x -in. recommended trapezoidal.

Height: ¾-in. to 1⅞-in., 1¼-in. to 1½-in. recommended.

13.12. Shroud Ring

This is recommended to give structural strength to the prongs or ends of the cap. The face of the ring may rest directly on the tray floor, or it is recommended to have three short legs of ¼-in. for clean service. For all materials the skirt clearance often used is ½ to 1-in., and for dirty service with suspended tarry materials upto l½ in. These legs allow fouling or sediment to be washed out of the tray, and also allow emergency cap action under extremely overloaded conditions at lower

efficiencies.

13.13. Tray Performance-Bubble Caps

A bubble cap tray must operate in dynamic balance, and the closer all conditions are to optimum, the better the performance. Evaluation of performance requires a mechanical interpretation of the relationship of the tray components as they operate under a given set of conditions. This evaluation includes the determination of:

1. Tray pressure drop
    a. Slot opening
    b. Static and dynamic slot seals
    c. Liquid height over weir
    d. Liquid gradient across tray

2. Downcomer conditions
    a. Liquid height
    b. Liquid residence time
    c. Liquid flow over weir into downcomer

3. Vapor distribution

4. Entrainment

5. Tray efficiency

The evaluation is made in terms of pressure drops (static and friction) through the tray system. Figure 13-1Figure 13-2Figure 13-3 and Figure 13-4 present the tray action diagrammatically.

An understanding of the action of the bubble cap tray is important in deciding upon the acceptability of a particular design. The passage of vapor through the caps, and liquid across the tray is made complicate by fluid actions associated with the mechanical configuration, and with the relative velocities of the fluids at various points on the tray. Quantitative considerations will be given in more detail in later paragraphs, however, qualitative interpretation is extremely valuable. The following descriptions are presented for this purpose.

13.13.1. Tray Capacity Related to Vapor–Liquid Loads

Figure 13-39 presents a generalized representation in the form useful for specific tray capacity analysis. Instead of plotting actual vapor load versus liquid load, actual vapor load per cap (here the cap row relative to inlet or outlet of tray is significant) versus the liquid load per inch or foot of outlet weir length is given.

Although each plot must be for a specific set of conditions, Figure 13-40 and Figure 13-41 are extremely valuable in analyzing the generalized action of a bubble tray.

For Figure 13-41, Bolles points out that the cap loads for inlet and outlet rows will be essentially balanced or “lined out” when the shaded areas are equal.

From Figure 13-39, the region of satisfactory tray operation is bounded by performance irregularities. Here vapor is flowing through all the caps; the bubbling action is acceptable from an efficiency standpoint; entrainment is within design limits; there is no dumping (or back flow) of liquid down the risers, and no undesirable vapor jetting around the caps.

13.13.2. Tray Balance

A tray is in balance when it operates with acceptable efficiency under conditions at or very near those of design.

13.13.3. Tray Flexibility

A tray is flexible when it operates with acceptable efficiency under conditions which deviate significantly from those established for design. The usual changes affecting flexibility are vapor and/or liquid loading. A tray may operate down to 50% and up to 120% of vapor load, and down to 15% and up to 130% of liquid load and still be efficient. Beyond these points its efficiency may fall off, and this indicated the flexible limits of the tray.

13.13.4. Tray Stability

A tray is stable when it can operate with acceptable efficiencies under conditions that fluctuate, pulse, or surge, developing unsteady conditions. This type of operation is difficult to anticipate in design, and most trays will not operate long without showing loss in efficiency.

13.13.5. Flooding

A bubble tray tower floods when the froth and foam in the downcomer back up to the tray above, and begin accumulating on this tray. The downcomer then contains a mixture that is of lower density than the clear liquid, its capacity becomes limited, disengagement is reduced, and the level rises in the downcomer. This level finally extends onto the tray above, and will progress to the point of filling

the column if not detected, and if the liquid and vapor loads are not reduced. Flooding is generally associated with high liquid load over a rather wide range of vapor rates. The foaming tendencies of the liquid influence this action on the tray. The design condition for height of clear liquid in the downcomer for flooding is usually set at 0.60 to 0.80 of ($S_t + h_w$). See Figure 13-1.

13.13.6. Pulsing

A bubble tray pulses when the vapor rate is low and unsteady, when the slot opening is small (usually less than ½ in.), and when the liquid dynamic seal is low. If vapor enters the caps irregularly, the liquid pulses or surges, even to the point of dumping or backflowing liquid down the risers. The best cure is a steady vapor rate and good slot opening to allow for reasonable upsets.

13.13.7. Blowing

A bubble tray blows when the vapor rate is extremely high, regardless of the liquid rate, causing large vapor streams or continuous bubbles to be blown through the liquid. The efficiency and contact are low; entrainment is usually high. In this case low slot seals also contribute to the sensitivity of the tray to such action.

13.13.8. Coning

A bubble tray cones when the liquid seal over the slot is low and the vapor rate high enough to force the liquid completely away from the cap, thus bypassing the liquid entirely. Obviously, efficiency is unsatisfactory. The dynamic slot seals recommended in Table 13-7 will normally prevent such action.

Table 13-7 Suggested Slot Seals

Used by permission, Bolles, W. G., Petroleum Processing, Feb. thru May (1956), and Davies, J. A., Pet. Refiner V. 29, No. 93 (1950), Gulf Pub. Co.

Tower operating pressure	Static slot seal [15], in	Dynamic slot seal [5], in

	[15], in.	[5], in.
Vacuum, 30–2000 mm Hg. Abs	0–0.25	0.5–1.5
Atmospheric	0.5	1.0–2.0
50–100 psig	1.0	1.5–3.0
300 psig	1.5	2.0–4.0
500 psig	1.5	2.0–4.0

13.13.9. Entrainment

A bubble tray has high entrainment when mist and liquid particles carry up in the vapor from the liquid on one tray through the riser and cap on to the tray above. Bubble caps tend to entrain by jetting liquid–vapor mixtures high above the tray. Sufficient tray spacing must be available to prevent the quantity of material significantly affecting the efficiency of the system. The quantitative presentation of entrainment given in later paragraphs is aimed at calculating appropriate values.

13.14. Overdesign

Overdesign is often necessary in designing a tray, although caution must be exercised to prevent safety factors adding together and resulting in numbers which are totally unrealistic for performance. In other words, the magnitude, effect, and significance of over-capacity figures must be continuously monitored as each is calculated. A factor of 10–15% on liquid and vapor rates is usually acceptable, but each should be checked for its effect on maximum cap vapor capacity and entrainment, and on liquid gradient and buildup in the downcomer.

13.15. Total Tray Pressure Drop

This is normally taken as the wet bubble cap pressure drop plus the “mean dynamic slot seal” in inches of clear or unaerated liquid on the tray.

Guide values for normal operations, drop per tray are:

13.16. Liquid Height Over Outlet Weir

For a straight (non-circular) weir, the head of liquid over its flat top is given by the modified Francis Weir relation (Figure 13-42; also see Figure 13-1):

(13-4)

Figure 13-40

Effect of vapor load on slot opening, slot load, and pressure drop.

Used by permission, Bolles, W. L. Pet. Processing; Feb.–May (1956).

Figure 13-41

Effect of liquid gradient on vapor distribution with 0.5 vapor distribution ratio.

Used by permission, Bolles, W. L. Pet. Processing; Feb.–May (1956).

Figure 13-42

Liquid height over straight weir.

The modifying factor F_w developed by Bolles [5] for restriction at the shell due to segmental downcomer application is determined from Figure 13-43.

Figure 13-43

Weir formula correction factor for segmental type weirs.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

When h_{ow} values exceed 1¾ to 2 in., consider special downcomers or down pipes to conserve cap area for high vapor loads.

Notched outlet weirs (usually 60° V-notch) are only used for low liquid flow rates, and the head over this type of weir with notches running full [13] is given by:

(13-5)

For notches not running full:

(13-5A)

where

h_{ow} = height of liquid crest over flat weir, in.
l_w = length of weir (straight), feet
L_g = liquid flow rate, gallons per minute, tray or tray section
n = depth of notches in weir, in.
= height of liquid above bottom of notch in weir, in.
h_n = depth of notches in weir, in.

(13-6)

where
d_w = diameter of circular weir, in.

13.17. Slot Opening

The slot opening is the vertical opening available for vapor flow during operation of the cap under a given set of conditions. It has been found to be essentially independent of surface tension, viscosity and depth of liquid over the cap. When the required opening exceeds that available, the vapor will either flow under the cap, thus creating greater pressure drop.

Caps with Rectangular Slots [60]

(13-7)

where
h_s = slot opening or pressure drop through slot, in. liquid
V = total vapor flow through tray, ft^3/sec
N_c = number of caps per tray

N_s = number of slots per cap
w_s = width of slot (rectangular) in.

Figure 13-44 presents a quick solution for this equation. Maximum slot capacity [5]:

(13-8)

where
A_s = total slot area per tray, ft^2
H_s = slot height. in.
V_m = maximum allowable vapor load per tray, ft^3/sec

Figure 13-44

Opening of rectangular-vertical slots for bell type bubble cap.

Used by permission, Stone and Webster Engineering Corp., Boston, Mass.

Caps with Trapezoidal Slots

A trial solution is involved in determining the slot opening for trapezoidal slots [5]. The relation for maximum capacity at full slot opening is:

(13-9)

where

R_s = ratio of top to bottom widths of trapezoid slot.

Figure 13-45 is useful in solving for the slot height as a percentage of vapor capacity and of full opening.

Figure 13-45

Trapezoidal slot generalized correlation.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

Whenever possible, slots should be designed to be 50–60% open to allow for pulses and surges in vapor flow.

13.18. Liquid Gradient Across Tray

The difference in height of liquid between the inlet and outlet sides of a tray is the liquid gradient. It is the result of frictional drag on the caps and internals, plus resistance created by the bubbling action. A tray with high liquid gradient may be operating inefficiently and at reduced capacity if the rows of caps covered by high liquid are not bubbling, thus forcing all the vapor through the rows of caps nearer the tray outlet where the liquid head is lower. Liquid gradient is one of the criteria which must be checked to assure proper understanding of a tray design and its performance.

The recommendation of Bolles [5] is based on the work of Davies [14, 15] and serves the average design adequately. It assumes an I.D. of bubble cap to I.D. of riser of 1.42 and this is close to the actual range

for 85% of the installations. Small deviations will be negligible. It must be remembered that the agreement between the several investigators is good [24, 38, 44], but still lacks a final solution to all situations. In general, calculated values should not be considered better than ±0.2 in.

The relation:

(13-10)

Solve for left side expression, and determine (Q/L_w) from Figure 13-46, or use Figure 13-47Figure 13-48Figure 13-49 and Figure 13-50, where

L_g = total liquid load in tray or tray section, gpm
$L_w = l_{fw}$ = total flow width across tray normal to flow, ft
C_d = liquid gradient factor
γ = ratio of distance between caps to cap diameter
= liquid gradient per row of caps, uncorrected, in.
h_l = depth of clear liquid on tray, in.
s = cap skirt clearance, in.
$L_w = l_{fw}$
$Q = L_g$

Figure 13-46

Modified liquid gradient factor chart for no hold-down bars.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

Figure 13-47

Liquid gradient chart - cap spacing 25% cap diameter.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

Figure 13-48

Liquid gradient chart - cap spacing 31.25% cap diameter.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

Figure 13-49

Liquid gradient chart - cap spacing 37.5% cap diameter.

Used by permission, Bolles, W. L., Pet. Processing, Feb.–May (1956).

Figure 13-50

Liquid gradient chart – cap spacing 50% cap diameter.

Used by permission, Bolles, W. L., Pet. Processing; Feb.–May (1956).

(13-11)

Some designers use Δ/5 to Δ/3 in place of Δ/2.

The charts of Figure 13-47Figure 13-48Figure 13-49 and Figure 13-50 were developed [5] from the modified Davies equation to simplify the solution of a tedious problem. The mean tray width is usually taken as average of weir length and column diameter. Special tray patterns may indicate another mean value.

The values of liquid gradient read from these charts are uncorrected for vapor flow. This correction is a multiplier read from Figure 13-51:

(13-12)

Figure 13-51

Correction of Liquid gradient for vapor load.

Used by permission, © The American Chemical Society, Davies, J. A., Ind. and Eng. Chem. V. 39, p. 774 (1947).

and V_o = V = vapor load for tray, ft^3/sec.

Although this method appears to be conservative for the average case, it is not strictly correct for towers with liquid flowing over the cap. Therefore it would be well to check the results of gradients over 1.0 in. by comparing with some of the other methods and with the tabulation of data of Reference 38.

Adjustments to the tray or caps are usually not considered unless the calculated gradient exceeds ½ to 1 in. liquid. Several schemes are in use:

1. Raise cap in inlet half of tower by one-fourth to one-half the calculated gradient, but not exceeding 1 in.
2. For large towers (usually over 8 ft in diameter) check the hydraulic gradient for sections of the tray normal to liquid flow, adjusting each section by not more than one-half the gradient.
3. Slope the trays downward from liquid inlet to outlet, with the total drop from inlet to outlet weir not exceeding one-half the calculated gradient.
4. Cascading the tray by using weirs as dams to divide the tray into steps, each step or section of the tray having no significant gradient from its inlet to outlet. This is usually only considered for trays 10 ft in dia. and larger, as it adds considerably to the cost of each tray.
5. More elaborate tests and adjustments can be made [5]. However, they are usually unnecessary except in the unusual cases of very high liquid loads and/or large columns.

In any case, the average head over the cap slots for the section should approximately equal the average head over the adjoining sections, and the inlet and outlets of the section should not be extreme, even though the average is acceptable. The object of a fairly uniform head over the slots should be kept in mind when reviewing the gradient adjustments.

13.19. Riser and Reversal Pressure Drop

The method proposed by Bolles fits the average design problem quite satisfactorily. However, for low pressure drop designs, as in vacuum towers, it may well require checking by the more detailed method of Dauphine [13].

A. Bolles' Design Method [5l]:

Solve for the combined riser, reversal, annulus, and slot pressure drop by:

(13-13)

where

h_{pc} = cap assembly pressure drop, including drop through riser, reversal, annulus, slots, in. liquid

A_r = total riser area per tray, ft^2

K_c = constant for Bolles bubble cap pressure drop equation, (Figure 13-52)

The constant, K_C, is obtained from Figure 13-52, noting that the annular area between riser and cap must always be larger than the riser area for K_C to be valid.

Figure 13-52

Bubble cap pressure drop constant (Bolles' Method).

Used by permission, Bolles, W. L. Pet. Processing, Feb.–May (1956).

B. Modified Dauphine Relations [5, 11]:

1. Riser pressure drop

(1) Reversal Area Greater than Riser Area

(13-14)

(2) Reversal Area Less than Riser Area

(13-15)

2. Reversal and annulus pressure drop

The reversal area is the area of the cylindrical vertical plane between the top of the riser and the underside of the bubble cap, through which the incoming vapor must pass. The vapor then moves into the annulus area between the inside diameter of the cap and the outside diameter of the riser before entering the slots in the cap.

The pressure drop due to reversal is independent of slot height and riser height, as long as riser height is greater than 2 in. and the cap slot height does not exceed the riser height [13]. When riser height is less than 2 in., the reversal pressure drop increases as long as the slot height is less than the riser height.

For riser height greater than 2.5 in.:

(13-16)

3. Dry cap pressure drop: rectangular slots.

The slot pressure drop through the dry cap increases nearly linearly with cap diameter.

(13-17)

4. Total dry cap pressure drop:

(13-18)

5. Wet cap pressure drop:

(13-19)

The correction factor, C_w, is obtained from Figure 13-53. Figure 13-53 applies to cap slots 1 in. through 2 in., and if slots are smaller (around ½ in.) the C_w factor increases by about 25% average (10–50%). The relation applies only to the pressure drop attributable to the conditions of liquid on the tray up to the top of the slots.

Figure 13-53

Correction for wet pressure d rop C_w, (after Dauphine);

Used by permission, Bolles, W. L. Pet. Processing, Feb.–May (1956), using data of Dauphine [13].

6. Total bubble cap pressure drop:

(13-20)

h_s from Equation 13-17 and Figure 13-44

13.20. Total Pressure Drop Through Tray

A. Bolles' Method

(13-21)

B. Dauphine Method

(13-22)

13.21. Downcomer Pressure Drop

The head loss in liquid flowing down the downcomer, under its underflow edge (and up over an inlet weir, if used) and onto the tray is important in determining the back up of liquid in the downcomer. There are many suggested relations representing this head loss.

A. Segmental Type Downcomer

1. Downcomer friction loss plus underflow loss:

(13-23)

or, (alternate):

(13-24)

2. Loss through inlet weir

When an inlet weir is used, the additional resistance to flow may be approximated by:

(13-25)

B. Circular or Pipe-Type Downcomer

These downcomers are suggested only where liquid flow is relatively small for the required tower diameter, allowing a maximum of space for bubble caps.

(13-26)

13.22. Liquid Height in Downcomer

The backup of clear liquid during flowing conditions must be determined in order to set the proper tray spacing. Tray spacing is usually set at twice the liquid height in the downcomer. This can be adjusted to suit the particular system conditions.

(13-27)

13.23. Downcomer Seal

The downcomer seal prevents vapor from the tray from bubbling into the downcomer (see Figure 13-1), whether the trays are bubble cap, valve or sieve types. If a seal weir is not included in tray design, then operation problems including flooding, weeping and unstable

performance, including pressure drop, are increased, particularly during the startup phase.

The major factors governing the required clearance under the downcomer (see Figure 13-1), and the distance between the bottom of the downcomer and the tray it is emptying onto are [190]: (a) downcomer sealing, (b) downcomer pressure drop, and (c) fouling and/or corrosive nature of the fluids. The smaller the clearance, the more stable will be the tray set-up due to the greater restriction of vapor flow into and up the empty liquid downcomer.

Referring to Figure 13-1, the weir height, h_w, must always be greater than the clearance under the downcomer, i.e., between bottom of downcomer and tray floor, h_{dc1}. Too smaller clearance can cause flooding of liquid in the downcomer. There are flow conditions where this condition may not be valid, therefore, the tray flow range from startup to overload should be examined before finalizing the physical details.

Some authors recommend clearance of ¼ in. to ½ in. less than the tray weir height, but always greater than ½ in. [190]. The bottom of the downcomer must be sealed below the operating liquid level on the tray. Due to tolerance in fabrication and tray level, it is customary to set the downcomer seal referenced to the weir height on the outlet side of the tray. Recommended seals, based on no inlet weir adjacent to the downcomer, and referenced as mentioned are given in Table 13-8.

Table 13-8 Downcomer Liquid Seal [15]

Used by permission, Davies, J. A., *Pet. Refiner*, v. 29, No. 13 (1950), Gulf Pub. Co., all rights reserved.

Tower Diameter, ft.	Seal, Outlet Weir height minus Distance Downcomer Off Tray Floor, in.
6 and below	0.5

7–12	1
13 and above	1.5

For trays with inlet weirs, seal values may be reduced if necessary for high flow conditions. A good tray design is centered about a 1.5-in. clearance distance between tray floor and bottom of downcomer edge.

13.24. Tray Spacing

Adequate tray spacing is important to proper operation during normal, as well as surging, foaming, and pulsing conditions. Because the downcomer is the area of direct connection to the tray above, the flooding of a tray carries to the tray above. To dampen the response, the tray must be adequately sealed at the downcomer and the spacing between trays must be approximately twice the backup height of liquid in the downcomer. Thus for normal design:

where

S_t = tray spacing, in

H_d = height of liquid in downcomer, in

Once foam or froth in the downcomer backs up to the tray above, it tends to be re-entrained in the overflowing liquid, making it apparently lighter, and accentuating this height of liquid-foam mixture in the downcomer. The downcomer must be adequate to separate and disengage this mixture, allowing clear liquid (fairly free of bubbles) to flow under the downcomer seal.

A tray inlet weir tends to ensure sealing of the downcomer, preventing the bubbling caps from discharging a mixture into the downcomer. The residence time in the downcomers is another criterion of adequate

tray spacing. Table 13-9 provides guidelines for selection of tray spacing. The recommended downcomer velocities for different liquids are:

Minimally foaming liquids:	0.39 – 0.69 ft/sec. (0.12–0.21 m/s)
Liquids with medium foaming tendency	0.29 – 0.59 ft/sec. (0.09–0.18 m/s)
Highly foaming liquids	0.19 – 0.59 ft/sec. (0.06–0.09 m/s)

Table 13-9 Guidelines for Selection of Tray Spacing

Source: Mukherjee, S., "Tray Column Design: Keep Control of the Details", Chem. Eng., pp 52–58, September 2005.

Description	Tray spacing, mm	Comments
Column diameters larger than 3,000 mm	>600	The tray support beams restrict crawling space available; hence the large tray spacing
Column diameters between 1,200 mm and 3,000 mm	600	This spacing is sufficiently wide to allow a worker to freely crawl between trays
Column diameters between 750 mm and 1,200 mm	450	Crawling between the trays is seldom required, because the worker can reach the column wall from the tray manways
Fouling and corrosive foaming tendency	>600	Frequent maintenance is expected
Systems with a high foaming tendency	At least 450 mm, but preferably 600 mm or higher	Required to avoid premature flooding
Columns operating in spray regime	At least 450 mm, but preferably 600 mm or higher	Required to avoid excessive entrainment
Columns		Lower tray spacing restricts allowable

Columns operating in froth regime	<450	Lower tray spacing restricts allowable vapor velocity, thereby promoting froth-regime operation

13.25. Residence Time in Downcomers

To provide reasonably adequate time for disengagement of foam and froth from the liquid in the downcomer, the total downcomer volume is checked against a minimum allowable average residence time of 5 seconds.

For various foaming characteristics of the liquid in the system, Kister [190] reports a recommendation of W. L Bolles based on course lectures at the University of New South Wales and the University of Sydney, as follows:

Foaming Tendency	Example	Residence Time, seconds
Low	Low MW hydrocarbons, and alcohols	3
Medium	Medium MW hydrocarbons	4
High	Mineral oil absorbers	5
Very High	Amines, glycols	7

Table 13-10 gives suggested downcomer clear liquid velocities based on relative foaming characteristics of the fluid on the tray at tray conditions.

Table 13-10 Suggested Downcomer Velocities

Used by permission, Pet. Processing, Bolles, W. L., Feb. thru may (1956).

Approximate tray spacing, in.	System foaming characteristics allowable clear liquid Velocities, ft/sec		
	High	Medium	Low
18	0.15–0.2	0.35–0.42	0.45–0.52

24	0.25–0.32	0.48–0.52	0.55–0.60
30	0.30–0.35	0.48–0.52	0.65–0.70
Typical representative system	Amine, glycerine	Oil systems	Gasoline, light hydrocarbons

13.26. Liquid Entrainment from Bubble Cap Trays

The work of Simkin, Strand, and Olney [64] summarizes most of the work of other investigators, and can be used for estimation of probable entrainment from bubble cap trays as shown in Figure 13-54. It is recommended that the liquid entrainment for design be limited to 0.10 mols/mol dry vapor.

Figure 13-54

Correlation of entrainment for bubble caps.

Used by permission, Bolles, W. L. Pet. Processing, Feb.–May (1956), using data of Simkin *et al.* [64].

Eduljee's [19] correlation of literature data appears to offer a route to evaluating the effect of entrainment on tray spacing and efficiency. It is suggested as another check on other methods. Figure 13-55 may be used as recommended:

1. Assume or establish a level of acceptable entrainment, such as 10 mols liquid/100 mols vapor.
2. Determine effect on efficiency by Colburn's relation (see Efficiency section).
3. Calculate total entrainment as pounds of liquid per hour, based on total vapor flow in tower.
4. Determine area of tray above caps (equals tower area minus area of two downcomers for cross-flow tower).

5. Calculate liquid entrainment, , as pounds of liquid per square foot of net tray area, equals (Step 3)/(Step 4), lbs/hr (ft^2)

6. Calculate vapor velocity, v_f, based on area of (Step 4), ft/sec

7. Calculate density factor,

8. Calculate

9. From Figure 13-55, read Log W_e*

10. Calculate S″ from

11. Assume a foam height, h_f, of approximately twice the height of the dynamic tray seal, in. This agrees with several investigators for medium foaming systems.

12. Minimum tray spacing , in.

where

v_f = vapor velocity based on free area above caps (not including two downcomers), ft/sec

W_e* = entrainment corrected for liquid properties and plate spacing

W_e = liquid entrainment mass velocity, lbs entrainment per minute per/ft^2, based on net tray area

S′ = effective tray spacing, distance between top of foam, froth, or bubbles, and tray above, in.

S″ = clear height above foam or froth (equals tray spacing minus foam height above tray floor), ft

μ = viscosity of liquid, centipoise

σ = surface tension of liquid, dynes/ cm

W_e' = entrainment (based on assumed allowance) lbs liquid/ (ft^2 free plate area) (hr)

h_f = height of top of foam above tray floor, in.

Figure 13-55

Eduljee's entrainment correlation for bubble caps.

Used by permission, Eduljee, H. E. British Chemical Engineer, Step. (1958).

Free Height in Downcomer

(13-28)

Slot Seal

The static slot seal is the fixed distance between the top of the outlet weir and top of the bubble cap slots.

The actual operating or dynamic slot seal is more indicative of conditions pertaining to the tray in operation and is [5]:

(13-29)

Note that this seal varies across the tray, although the tray design must be such as to make the value of h_{ds} nearly the same for each row of caps.

In order to ensure good efficiencies and yet a definite seal consistent with allowable pressure drops, suggested values for h_{ds} are modified from the references and shown in Table 13-7.

Bolles [5] recommends dynamic slot seals on bubble caps to check against calculated values. See Table 13-7.

Inlet Weir

The inlet weir, see Figure 13-30, for any tray, i.e., bubble cap, valve or sieve, is important in ensuring a seal on the inlet downcomer as well as maintaining a more uniform liquid level across the flowing tray. The recessed seal pan, Figure 13-1, provides the same benefits plus it reduces sieve tray leakage on the inlet side of the tray due to lower immediate liquid head increase usually occurring at the tray weir. It is necessary to drill weep holes for drainage of the tray at shutdown in the blocked-off inlet weir area, but limit the number and size of holes to avoid excessive weep-drainage during tray operation.

13.27. Bottom Tray Seal Pan

This seal acts like a typical downcomer seal from the tray above, and should be dimensioned approximately the same, except:

1. To avoid liquid backup in the downcomer, provide a downcomer height that is about 1.5 times the selected tray spacing in the column.
2. To ensure non-surging liquid flow under the downcomer, use a clearance, h_{dcl}, of at least 3 times the design for the other trays, or a minimum of 2 in. to 4 in.
3. Enlarge the clearance between the downcomer face and the inlet weir (or equivalent), (see Figure 13-30 or 13-1) to 1.5 to 2 times the dimensions used for the other trays.
4. Provide drainage holes in seal pan to allow adequate drainage, flushing and cleaning, but not too many (number), which would prevent liquid backup sufficient to maintain a seal on the tray.

13.28. Throw Over Outlet Segmental Weir

To ensure unobstructed vapor passage above the froth and liquid in the downcomer from a tray, the liquid mixture must not throw against the shell wall. The distance of throw over the weir is given by Reference 5. See Figure 13-1.

(13-30)

(13-31)

For center downcomers as in a two-pass design, the throw must be conservatively less than a distance that would cause the opposing streams entering the same downcomer to interfere with each other. Sometimes the installation of a splash baffle will help avoid conditions leading to flooding and loss of tray efficiency.

13.29. Vapor Distribution

The vapor distribution approximation:

(13-32)

is an indication of the uniformity of vapor flow through the caps on the inlet side of the tray to those on the outlet side, and the tendency of inlet caps to stop bubbling. Davies [14,15] recommends values for the ratio of 0.4, not to exceed 0.6; Bolles recommends 0.5. Only at values around 0.1 is essentially uniform vapor flow maintained through all the caps. As the R_v ratio increases, a smaller percentage of the vapor flows through the inlet tray caps and a larger percentage is

shifted to the outlet caps. This cross-flow of vapor increases the effect of Δ on the tray and accentuates the problem. Unfortunately it is often overlooked in many tray examinations. When the ratio reaches the value equal to the cap drop at full slot capacity (usually 1.75 to 2.5 in.), liquid will flow or dump back down the risers at the inlet row of caps. This is definitely an improper tray operation, predicting markedly reduced tray efficiencies.

For stepped caps (tray level, risers of different heights) or cascade tray (tray consisting of two or more levels) care must be taken to analyze all the conditions associated with level changes on the tray. Reference 5 discusses this in some detail.

Example 13-1: Bubble Cap Tray Design

Check the trays for a finishing tower operating under vacuum of 75 mm Hg at the top. The estimated pressure drop for the twenty trays should not exceed 50–60 mm Hg. Fifteen trays are in the rectifying section, five in the stripping section.

Following the suggested form, and starting, with the standard tray design existing in the unit, the calculations will be made to check this tray, making modifications if necessary. The 6 ft 0 in. diameter tower was designed originally for a significantly different load, but is to be considered for the new service. The tray features are outlined in Table 13-5. It becomes obvious that the tray is too large for the requirements but should perform reasonably well. The weakest point in performance is the low slot velocity.

*Tray Design**

Tower application or service: Product Finishing

Tower Inside Diameter: 6 ft, 0 in.

Tray Type: Cross Flow

Tray Spacing, 24 in.; Type outlet weir: End

Note: Tray spacing was set when the 6 ft, 0 in. diameter as determined.

No. downcomers/tray 1,. Located: End

Cap Data:

1. Cap I.D, I.D., 3⅞ in., Spacing: 5½ in. Δ 60° centers
2. Total height, 3 in.
3. No./tray, N_c, 129
4. Slots: No., N_s, 50
5. Height, H_s, 1½ in.
6. Width, w_s, ⅛ in.
7. Skirt Height, s, ¼ in.
8. Shroud Ring height, h_{sr}, ¼ in.
9. Height of inside surface of cap above tray, 3.94 in.
10. Riser I.D., 2.68 in.
11. Riser height above tray floor, 3 in.

Areas:

12. Riser inside crosssectional, a_r, 5.43 in.^{2} per riser
13. Total riser inside crosssect. area/tray, A_r, 4.95 ft^2
14. Riser outside crosssectional area a_{ro}, 5.94 in^2 per riser. Riser is 2¾-in O.D.; $\pi\ 2.75^2/4 = 5.94$ in^2
15. Cap inside crosssectional area a_c, 11.79 in^2 per cap. Cap is 3⅞-in. I.D.; $\pi\ (3⅞)^2/4 = 11.79$ in^2
16. Total cap inside crosssectional area, A_c, 10.53 ft^2
17. Annular area per cap, a_a, in.^{2}, $(11.79 - 5.94) = 5.85$ in^2
18. Total annular area per tray, A_a, 5.24 ft^2

19. Reversal area per cap, a_r', in.^{2} = π(2.69) (3.94 − 3.0) = 7.95 in^2. d = (2.75 + 2.63)/2 = 2.69 in.
20. Total reversal area, per tray, A'_r, ft^2 (129/144) (7.95) = 7.12 in^2
21. Slot area per cap, a_s, (50) (⅛) (1.5) = 9.39 in.^{2}
22. Total slot area per tray, A_s, 8.40 ft^2

Tray Details

23. Length of outlet overflow weir, l_w, 4.0 ft
24. Height of weir (weir setting) above tray floor, h_w, 2.5 in.
25. Inlet weir (downcomer side) length (if used), 4.0 ft
26. Inlet weir height above tray floor, 3 in.
27. Height of top of cap slots above tray floor, 2 in.
28. Static slot submergence or static slot seal (2.5 − 2.0), h_{ss}, 0.5 in.
29. Height of bottom of downcomer above tray floor, 2¾ in.
30. Downcomer flow areas: (a) Between downcomer and tower shell, 0.886 ft^2
31. (b) Between bottom downcomer and tray floor, 0.710 ft^2
32. (c) Between downcomer and inlet weir, 0.740 ft^2
33. Riser slot seal, (3.0 − 2), 1.0 in.

Tray Operations Summary and Pressure Drop		**Top**
A.	Tray number	20
B.	Operating pressure, mm. Hg	75
C.	Operating temperature, °F	60
D.	Vapor flow, lbs/hr	6,565
E.	Vapor volume, ft^3/sec @ operating conditions, V	132.2
F.	Vapor density, lbs/ft^3 @ operating conditions	0.0138

G.	Liquid flows, gallons/minute, L_g	3.74
H.	Liquid flow, lbs/hr.,	1,515
I.	Liquid flow, ft^3/sec @ operating conditions	0.00834
J.	Liquid density, lbs/ft^3 @ operating conditions	50.5
K.	Superficial vapor velocity, based on Tower I.D., ft/sec, 132.2/28.28	4.67
L.	Vapor velocity based on cap area between inlet and outlet weirs, ft/sec 132/[28.28 – 2(2.12)]	5.49
M.	Volume of downcomer: Area top segment, *Perry's Handbk*. 3rd Ed. Pg. 32. h/D = 9 in./72 = 0.1276, A = 0.05799 $(6)^2$ = 2.08 ft^2 lower taper, use h @ ⅔ of vert. taper for estimate. 8/72 = 0.111, A = 0.04763 $(6)^2$ = 1.71 ft^2, Volume = (2.08) (0.5) + (1.71) (21/12) = 4.04 ft^3	4.04
N.	Liquid residence time in downcomer, seconds, (4.04)/0.00834 = 485	485
O.	Throw over downcomer weir (sideflow), inches	1.17
P.	Throw over downcomer weir (center flow), min. =	–
Q.	Tray layout, actual downcomer width, in. Taper downcomer has 6 in. vertical dimension at 9 in. wide. Tapers to 5½ in., 24 in. below tray	9 5½
R.	Slot velocity: maximum = $3.4/(\rho_G)^{½}$ ft/sec	29
S.	Slot velocity: maximum = $12.1/(\rho_G)^{½}$ = $12.1/(0.0138)^{½}$ and $12.1/(0.01735)^{½}$, ft/sec	103.1
T.	Slot velocity: superficial, $u_o = V/A_s$ = 132.2/8.40 and 105/8.4 ft/sec	15.7

Pressure Drop, Inches Liquid on Tray	Top	Bottom
a. Height of liquid over weir (straight weir) $l_w/D = 4/6 = 0.667$ Read $F_w = 1.018$ from Figure 13-43 $h_{ow} = 0.092\ (1.018)\ (3.74/4)^{2/3}$ Use ¼ in. – V-notched weir, 2.5 in. from tray floor to bottom of notch. This is necessary because of low liquid flow	0.0989	0.0989
b. Static submergence, h_{ss}, in.	0.5	0.5
c. Caps Modified Dauphine and Cicalese, [11, 13] dry cap basis.		
1. Riser pressure drop, reversal area greater than riser area	0.0633	0.0462

2. Reversal and annulus pressure drop Riser height > 2.5 in.	0.045	0.0322

3. Rectangular slot dry pressure drop	0.0308	0.0231

4. Total dry cap pressure drop	0.1391	0.1015

5. Wet cap pressure drop From Figure 13-53, C_w = 0.16	0.87	0.847
6. Check maximum pressure drop through wet caps: , in. Since h_c is less than h_c max., cap is O.K. and not blowing under shroud ring Bolle's recommendation	1.8	1.3

7. Riser, reversal, annulus pressure drop From Figure 13-52, K_c = 0.598	0.118	0.0861

8. Slot pressure drop, Rectangular slots	0.626	0.566
d. Liquid Gradient Mean tray width = (4 + 6)/2 = 5 ft GPM/ft mean tray width = 3.74/5 = 0.75 Assumed mean liquid depth, h_1 = 2.5 + 0.0989 + 0.1 Uncorrected / row caps = approx. 0.02 in. C_v, from Figure 13-51= estimated 0.55 (off chart) No. cap rows = 11 Corrected = (0.02) (0.548) (11) = 0.1206 inches , inches (essentially negligible in this case)	0.12 0.06	0.12 0.06
e. Total pressure drop per tray, in. liquid Modified Dauphine Bolles	1.528 1.502	1.505 1.310
f. Pressure drop for 15 trays in rectifying section Modified Dauphine, 15 (1.528) = 22.9 in. liquid = 34.2 mm Hg Bolles, 15 (1.502) = 22.4 in. liquid = 33.4 mm Hg Pressure drop for 5 trays in stripping section Modified Dauphine, 5 (1.505) = 7.52 in. liquid = 11.1 mm Hg Bolles, 5 (1.42) = 6.56 in. liquid = 9.7 mm	34.2 mm 33.4 mm 45.3 mm	11.1 mm 7 mm

Total pressure drop for 20 trays Modified Dauphine Bolles	43.1 mm	
g. Height liquid in downcomer		

1. Segmental, underflow plus friction

2. Segmental, upflow when inlet weir is used

3. Total segmental loss, h_d

4. Circular downspout

5. Liquid height in downcomer

6. Free height in downcomer **7. throw over weir**	0.000077 Neg. 0.000077 4.58 21.69 1.17	0.000077 Neg. 0.000077 4.56 21.71 1.17
h. Vapor distribution ratio	0.138	0.141
i. Slot seal Dynamic,	0.65	0.65

Liquid Velocity in Downcomer

Minimum crosssection area of downcomer = 0.886 ft^2

Liquid rate = 0.00834 ft^3/sec

Velocity = 0.00834/0.886 = 0.00942 ft/sec

This is very low and confirms that there should be ample disengaging capacity in the downcomers. The downcomers are larger than they should be.

Slot Velocity

The results of lines R, S, and T indicate that the vapor velocity through the cap slots is lower than desirable for good bubbling.

Slot opening

The slot opening, h_s, given in line c8 is only slightly lower than the

normal design of 50–60% of H_s, or 0.75 in. to 0.90 in.

Vapor Distribution Ratio

The values of line (h) are quite in line with good vapor flow through all the caps. This is as would be expected, since the hydraulic gradient is low; too low to require any compensation.

Liquid Entrainment

$v_f = 132.2/[28.28 - 2(2.12)]. = 5.5$ ft/sec

Reading Figure 13-54 for 27 dynes/cm surface tension

$W_e/h_{ow} + h_{ss} + h_s = 0.026$

$W_e = (.026)(0.098 + 0.5 + 0.626) = 0.0317$ 1b/min ft^2

Entrainment = (0.0317) [28.28 − 2(2.12)] = 0.764 1b/min

Entrainment ratio = 0.7641 (6565/60) = .00698

This value of entrainment is negligible. For a new column design, this would indicate that the tower was too large, and a smaller shell should be considered.

Conclusion

This is not a good tray design, but it should operate. However, a reduced efficiency is to be expected due to low vapor velocities.

Because the liquid flow is low, ¼-in. v-notched weirs should be used

to ensure uniform flow and level across the tray. The bottom of the notches should be 2.5 in. above the tray floor.

13.30. Sieve Trays with Downcomers

The performance analysis of these trays is fairly similar to bubble caps, but more so to valve trays, because the tray has the same basic mechanical features. The difference is that bubble caps and valves are replaced by perforations or holes in the tray to allow into, the gas entry to the liquid on the tray. Figures 13-5A and Figure 13-56 and Figure 13-57 represent the general construction of a sieve tray.

Figure 13-56

Sieve tray with downcomers, tower assembly.

Used by permission, Hendrick Mfg. Co.

Figure 13-57

Sieve tray with downcomers, tray components.

Used by permission, Hendrick Mfg. Co.

Sieve trays have been used in both clean and fouling service, including solutions containing suspended particles. The bubbling action seems to wash the solids down from tray to tray provided there

are no corners or “dead” spots. Sieve trays are preferable for applications operated at 50–100% capacity as long as surges between significantly different capacities are not expected. When operating within design range, the efficiency of these trays is better than for bubble cap trays, although without specific test data it is difficult to take advantage of this feature in safety.

In some sieves the capacity is 1.5 to as much as 3 times that of a bubble cap tray. The “type tray” guide proposed by Huang and Hodson [30] serves to identify the major breaks in type of tray design (Figure 13-58). In the region between types, the selection is not sharp and the design should be evaluated based on other criteria.

Figure 13-58

Selection guide, perforated trays with downcomers.

Used by permission, Chen-Jung and J. R. Hudson, Petroleum Refiner, V. 37, p. 104 (1958), Gulf Publishing Co., All rights reserved.

Various aspects of sieve tray performance have been studied [30, 31, 33, 36, 41, 42, 45, 71, 78] and several design methods have been recommended [30, 31, 41, 42, 189, 197]. The following composite method has given good performance in operating towers, and is based on satisfying the three critical capacity features; entrainment, flooding, and weeping.

This type of tray seems to produce fewer jets of liquid froth than a bubble cap tray. The entrainment from the surface of the bubbling

liquid-froth mixture is less (about ⅓) than a bubble cap tray for the same superficial tower velocity and tray spacing. Generally, the trays will flood before capacity reaches the limit a set by entrainment.

The proprietary “Linde Tray”, Figure 13-5C, is a proven tray design used for new installations [198], and also for improving the performance of existing distillation columns by replacing older and possibly less efficient trays. One of the advantages of this tray type is its capability of being installed at tray spacings as low as 9 to 10 in. and frequently at 12 in. The tray efficiency varies with distillation system, but as a general guide, will be equal to that of a multipass tray.

A definition of terms, some more related to sieve trays than other types, is provided by Chase [192] (used by permission, Chem. Eng., July 31, 1967):

Cross flow: Liquid flowing across a plate (rather than straight down through the holes) so that it falls to the plate below through a channel at one side of the plate.

Liquid throw: The horizontal distance traveled by the liquid after flowing over a weir.

Dual flow: Both liquid and vapor pass through the perforations on the tray; there are no downcomers.

Radial flow: Liquid flowing radially from, or to, an inlet (or outlet) located at the center of the tray, to (or from) downcomers (or inlets) at the tray periphery.

Reverse flow: Liquid flowing from the inlet on one side of the tray (around a center baffle) reverses its direction at the other side of the tray, and flows back to the downcomer on the same side of the tray

where the inlet is.

Split flow: Liquid flow across the tray is split into two or more flow paths.

Double pass: A split-flow tray with two liquid flow paths on each tray. Each path handles half of the total liquid flow.

Blowing: A condition where the rising vapor punches holes through the liquid layer on a tray and usually carries large drops and slugs of liquid to the next tray.

Coning: A condition where the rising vapor pushes the liquid back from the top of the hole, and passes upward with poor liquid contact.

Dumping: A condition caused by low vapor rates where all of the liquid falls through some holes (rather than over the weir) to the tray below, and vapor rises through the remaining holes.

Raining: A condition similar to dumping (no liquid goes over the weir) except that, because of higher vapor rates, the liquid fall through the holes is more uniform.

Weeping: A condition occurring when the vapor rate is not large enough to hold all the liquid on the tray, so that part of the liquid flows over the outlet weir while the rest falls through the holes.

Flooding: A condition that gives rise to a sharp decline in tray efficiency and a sharp increase in pressure drop. Flooding is commonly due to either an excessive carryover of liquid to the next tray, or to an inability of the system to convey the liquid flow to the tray below.

Oscillation: A wave-type motion of the liquid on the tray, perpendicular to the normal direction of flow.

Seal point: The point at which a weeping condition changes to raining.

Injection regime: A condition in which the liquid above the plate is in the form of individual drops dispersed in the vapor; thus, there is virtually no mixing in the main bulk of the liquid.

Stable regime: The preferable hydrodynamic condition of the aerated liquid on a sieve tray. The aerated material exists as a stable froth; gas-liquid contact is good.

Turndown ratio: A term used by designers to denote ratio of minimum allowable to operating throughput.

Segmental downcomer: The channel for liquid flow formed by an enclosed segmental tray section.

F factor: The vapor kinetic energy parameter, often used as a correlating term for flooding velocity, foam density, *etc.*

Souders and Brown equation: $G/A = K[d_v (d_L - d_v)]^{1/2}$, where G/A = superficial vapor flow, lb/ (hr) (ft^2) , and d_v and d_L = vapor and liquid densities, lb/ft^3. See Equation 13-1.

13.31. Tower Diameter

The tower diameter may be calculated for first approximation by the Souders-Brown method; however, this has been found to be conservative, since it is based on no liquid entrainment occuring between trays. (Actually, some entrainment can be tolerated at negligible loss in efficiency or capacity.)

There are several approaches to column diameter design [65, 74] as

well as the proprietary techniques of major industrial and engineering designers. Some of these use the proprietary Fractionation Research Institute (FRI) methods which are only available on a membership basis and do not appear in the technical literature.

In general, a better first approximation and often a more economical tower diameter is determined using Figure 13-59 [33].

(13-33)

(13-34)

where

e_w = weight of liquid entrained/unit weight of vapor flowing in sieve tray column

σ = liquid surface tension, dynes/cm

v_c = vapor velocity based on column crosssection, ft/sec

S' = effective tray spacing, distance between top of foam and next plate above, in.

h_c = height of clear liquid in bubbling zone, in.

Figure 13-59

Sieve tray entrainment correction.

Used by permission, Hunt, C. D. A., Hanson, D.N., and C.R., Wilke, The American Institute of Chemical Engineers, Chemical Engineers Journal, V. I, p. 441 (1955), All rights reserved.

This is based on a frothed mixture density of 0.4 times that of the clear liquid on the tray, which has been found to be a reasonable average for several mixtures.

Entrainment values of 0.05 lbs liquid/lb vapor are usually acceptable, with 0.001 and 0.5 lb/lb being the extremes. The specific design dictates the tolerance on entrainment. From the calculated vapor velocity, v_c, the diameter of the column can be calculated using:

(13-35)

Entrainment does not usually become a problem until the tray is operating at 85–100% of the flooding condition. Figure 13-59 is convenient for solving for e_w.

13.32. Tray Spacing

Tray spacing can usually be about 6 in. less than for a corresponding bubble tray. Sieve trays are in operatoin with spacings of 9 in. and up to 30 in., the latter being necessary for high vacuum service. Spacing of 12–16 in. is common.

Minimum spacing is the same as recommended for bubble cap trays, i.e., $S_t = 2\ H_d$.

13.33. Downcomer

Downcomers are designed for the same conditions as bubble tray towers.

Biddulph, Thomas, and Burton [209] studied the effects of downcomer designs, i.e., chordal or segmental, circular downpipe,

low liquid flows, sloped downcomer (good for disengaging foam/bubbles, etc.) and envelope type, for use with sieve trays. They developed a modification of the segmental style by installing a downcomer weir on the tray floor inside the weir outlet (see Figures 13-60 A and 13-60 B). This replaces the usual weir, which is placed outside of the outlet of the downcomer. Note that it runs for only about 75% of the chordal length of the downcomer width. The authors state that this still provides a liquid seal all along the inlet, but does provide space at the ends to exert a positive influence on the tray liquid flow pattern. For the segmental downcomer:

1. Mechanism 1 of Figure 13-60 B [209] is dominant when the underflow clearance at a given liquid rate is increased, the underflow velocity decreases and the severity of recirculation decreases.
2. Mechanism 2 of Figure 13-60 B becomes apparent when the flow recirculation on the tray increases with increasing underflow clearance. The curvature of the column wall influences the movement of the liquid toward the center. High underflow clearance does not even out maldistribution due to backup where the irregular flow pattern enters into the tray below. This allows flow separation to occur on the downcomer floor, and leads to enhanced retrograde flow.

Figure 13-60

Modification of downcomer weir at tray floor outlet; (A) downcomer weir; (B) mechanism of flow separation.

Used by permission, Biddulph, M. W. et al., The American Unstitute of Chemical Engineers, Chem. Eng. Prog., V. 89, No. 2, p. 56 (1993), All rights reserved.

Biddulph [209] *et al.* summarizes “rules of thumb” that have been expressed elsewhere in the literature for downcomer sizing (used by permission of Chem. Eng. Prog. V. 89, No. 12, 1993).

Rules of thumb that have developed out of many years of industrial experience relating to downcomer sizing include:

1. Set the velocity of the unaerated liquid under the downcomer to 1.6 ft/sec (0.5 m/sec).
2. Let the liquid velocity under the downcomer equal the liquid velocity on the tray to give a smooth entry [237].
3. Hold the head loss due to the underflow clearance, h_{udc}, to no more than 1.0–1.5 in. of hot liquid [117].
4. Allow at least 3 sec residence time in the downcomer for disengagement of vapor for a nonfoaming system, and 6 sec for a foaming system [238].

Figure 13-61 illustrates a typical sieve tray capacity chart. Entrainment by jet flooding or limitation by downcomer flooding are two of the main capacity limiting factors. The liquid backup in the downcomer must balance the pressure drop across the tray, with the process balance [209]:

(13-36)

where

h_{fd} = downcomer backup, in.,

h_{WT} = wet tray head loss, in.

h_{li} = clear liquid head at the inlet to tray, in.

h_{udc} = head loss due to underflow clearance, in.

h_n = head in the back of downcomer, in. (usually negligible except at high liquid load)
α_d = mean aeration factor of froth, dimensionless (see Figure 13-64)

Figure 13-61

Sieve tray capacity chart.

Used by permission, Biddulph, M. W. et al., The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 89, No. 2, p. 56 (1993), All rights reserved.

13.34. Hole Size and Spacing

Most of the literature has presented data for trays with holes of ⅛-in. through ¼-in. diameter. The work of Hunt *et al.* [33] includes ½ in. holes. Some commercial units have used ¾ in and 1-in. holes, although these sizes should be used with caution when adequate data are not available. The recommended hole size for the average clean service is -in. based on present published data. Holes of ½ in. may be used for any service including fouling and fluids containing solids with no loss in efficiency. Holes of ⅛ in. dia. are often used in vacuum service.

Holes spaced closer than twice the hole diameter lead to unstable operation. The recommended spacing is $2.5d_o$ to $5d_o$ with $3.8d_o$ being preferable [42]. Holes are usually, placed on 60° equilateral triangular pitch with the liquid flowing normally to the rows. Holes should not be greater than 2.5-3 in. apart for effective tray action.

The percentage hole area in a tray varies according to the needs of the design; the usual range is 4–15% of the total tower crosssection. Experience has indicated that this is a questionable basis, and it is clearer to refer areas to the active bubbling section of the tray, provided liquid cannot by-pass this area. Thus, rather arbitrarily, but referenced to test literature, the effective tray action area might be the area enclosed by encircling the perforated hole area a distance 2–3 in. from the periphery holes. On this basis, the hole area would be 6–25% with a usual value range of 7–16% with about 10% being preferred.

13.35. Tray Hydraulics

Figure 13-62 illustrates a typical pressure drop diagram for a sieve tray. Note that the figure is for liquid flowing with gas countercurrent. For dry tray, only gas is flowing and no liquid, and the pressure drop is a function of the orifice coefficient. For wet tray pressure drop, gas and liquid are both flowing and the pressure drop is a function of clear liquid head, head over the weir, hydraulic gradient, residual pressure drop, foam density and height, aeration and twophase regime factors, and bubbling frequency [192]. The pressure drop associated with the downcomer is a function of liquid backup, foam density and aeration factor, and liquid throw at the outlet weir [192]. See Figure 13-40, which relates similar factors for bubble cap trays, as well as valve trays. Figure 13-63 [192] presents a typical performance diagram of the operating features of a sieve tray.

Figure 13-62

Typical operating curve of sieve trays with downcomers. Note modes of operation.

Used by Permission, Chase, J. D., Chem. Eng., July 31, p 105, (1969). Also see Klein [201], Figure 13-86.

Figure 13-63

Performance diagram of sieve trays (note article reference No. 18).

Used by permission, Chase, J. D., Chem. Eng., July 31, p 105 (1969)

Figure 13-64

Aeration factor, sieve trays.

Used by permission, Smith, B. D., Design of Equilibrium Stage Processes, Chapter 15, by J.R. Fair, McGraw-Hill Book Co. (1963), All rights reserved.

13.36. Height of Liquid Over Outlet Weir, h_{ow}

This may be calculated as recommended for bubble cap trays. Minimum weir height is 0.5-in., with 1–3 in. preferred. See Figure 13-5 A.

13.37. Hydraulic Gradient, Δ

Tests have indicated that the hydraulic gradient is negligible or very small for most tray designs. Usual design practice is to omit its effect unless the value of Δ is expected to be greater than 0.75 in. If hydraulic gradient is appreciable, then the holes nearer to the tray inlet

(liquid) will tend to weep before those nearer the tray outlet.

This creates the same type of cross-flow and improper distribution as was discussed for bubble cap tray operation. The recommendation of Hughmark and O'Connell [131] includes corrections to the friction factor of Klein [39]. For stable tray operation, the hydraulic gradient should be less than one-half the dry tray pressure drop. For conditions of high weir height and high $v_o(\rho_v)^{1/2}$ the greater the friction factor affecting the hydraulic gradient [25]. Also, the greater the liquid flow the higher the pressure drop and gradient.

For the tray liquid to move from inlet to outlet of the tray, there must be a liquid flow gradient on the tray in that direction. (see Figure 13-5 A). The sieve tray usually has less problems with liquid gradient than bubble cap or valve trays, the general guide for avoiding gradient problems (good tray stability) is similar to bubble cap design [193]:

(13-37)

(13-38)

Use velocity of aerated mass same as for clear liquid.

(13-39)

(13-40)

where

l_{fw} = total flow width across tray, normal to flow, ft., For this equation, use arithmetic average between tower diameter, D, and weir length, l_w.

h'_f = height of froth (aerated mass) above tray floor, in., estimated from discussion under "Total Wet Tray Pressure Drop" (see Figure 13-64)

f = friction factor for froth cross-flow

l_w' = length of flow path, ft

g = acceleration of gravity, ft/sec-sec

h_l = equivalent height of clear liquid on tray, in

h_{lo} = height of clear liquid at overflow weir, in

h_{li} = height of clear liquid on inlet side of tray, in

h_w = height of weir above tray floor, in

h_h = head loss due to vapor flow through perforations, in. liquid

ρ_1 = density of clear liquid, lb/ft^3

μ_1 = viscosity of liquid, lb/ft sec

q = liquid flow rate, ft^3/sec

v_f = velocity of froth cross-flow, ft/sec

Figure 13-65 [193] is used to determine friction factor, f.

Figure 13-65

Friction factor for froth cross flow, sieve trays. (Note extrapolation by Ludwig, E. E).

Used by permission, Smith, B. D., Design of Equilibrium Stage Processes, Chapter 15, by J.R. Fair, McGraw-Hill Book Co. (1963), All rights reserved.

Reynolds No. Modulus:

(13-41)

The relationship between f and Re_h is given in Figure 13-65 and is recommended for design purposes. The velocity of the aerated mass is the same as for the clear liquid.

(13-42)

13.38. Dry Tray Pressure Drop

This is the drop occurring when the vapor passes through the holes on the tray. The relation below [25] correlates the data of several of the major investigators with a maximum deviation of less than 20% and an average deviation of 10%.

(13-43)

(13-44)

where

h_{dt} = pressure drop through dry perforated tray, inches liquid on tray

v_o = vapor velocity through perforated holes, ft/sec

β = fraction perforated hole area in perforated tray area only

C_o = orifice coefficient from Figure 13-66

Figure 13-66

Orifice coefficient for perforated trays.

Used by permission, Hughmark, G. A., and H. E. O'Connell, The American Institute of Chemical Engineers, Chem. Eng. prog., V. 53, p.127M (1957), All rights reserved.

Note that β is not the fraction of hole area in the active tray region, but is limited to the perforated section only.

13.39. Fair's Method [193]

This method calculates the dry tray pressure drop and allows for correcting the twophase flow effects at various entrainment ratios.

(13-45)

C_o is a function of the velocity of approach, hole diameter/tray-thickness ratio, Reynold's number through the hole, condition or design of the “lip” of the hole, and some other less prominent variables. The correlation for this concept for the orifice discharge

coefficient is from Liebson, *et al.* [42], see Figure 13-67. Use C_o from this figure in Equation 13-45.

Figure 13-67

Discharge coefficients for vapor flow, sieve trays.

Used by permission, Smith, B. D., Design of Equilibrium Stage Processes, Chapter 15, by J.R. Fair, McGraw-Hill Book Co. (1963); data from I. Leibson, R. E. Kelley, and L. A. Burlington, Petroleum Refiner, V. 36 (2) Feb. (1957), p. 127; V. 36 (3) p. 288 (1957). All rights reserved.

where

A_h = net perforated area of tray, ft^2

A_a = active or "bubbling" area of tray, generally, $(A_t - 2A_d)$, ft^2

A_d = downcomer area, crosssectional area for total liquid downflow, ft^2

A_t = total tower crosssections, area, ft^2

C_o = vapor discharge coefficient for dry tray

g = acceleration of gravity, 32.2 ft/sec^2

h_h = head loss due to vapor flow through perforations, in. liquid

v_o = vapor velocity through perforations, ft/sec

ρ_1 = clear liquid density, lb/ft^3

ρ_v = vapor density, lb/ft^3

13.40. Static Liquid Seal on Tray, or Submergence

Aeration of the liquid by gas bubbles reduces its density. The usual and somewhat conservative approach recommends that this aeration effect be neglected. Many successful towers have trays operating on

this design basis [45],

A.

(13-46)

$f = 1.0$

where f = aeration factor

h_{sl} = static liquid seal on sieve tray, in. liquid

B. A second and also successful method accounts to a certain extent for the aeration effect, based on test data from many references. This method is not quite as conservative when estimating total tower pressure. This follows the effective head concept of Hughmark *et al.* [31]. Effective head, h_e, is the sum of the hydrostatic head plus the head to form the bubbles and to force them through the aerated mixture. Figure 13-68 is the correlation for h_e plotted against submergence, h_{s1} [31]. See "Dynamic Liquid Seal."

Figure 13-68

Effective liquid head for sieve trays with downcomers.

Used by permission, Hughmark, G. A. and H. E. O'Connell, The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 53, p. 127M (1957). All rights reserved.

13.41. Dynamic Liquid Seal

When hydraulic gradient is a factor in tray design, the dynamic liquid seal should be used in place of h_{sl} for the determination of effective

head:

(13-47)

where

h_{dl} = dynamic liquid seal for sieve tray, in. liquid

h_e = effective liquid head taking aeration of liquid into account, in. liquid, from Figure 13-68

The aerated liquid pressure drop includes that generated by forming bubbles [193] due to surface tension effects. The equivalent height of clean liquid on the tray is given by [193]:

(13-48)

The term, h_1, represents the hydrostatic head on the tray, while (h_w + h_{ow}) is the liquid seal at the tray outlet weir, expressed as clear liquid. The factor, β, can be obtained from the upper curve in Figure 13-64 [193].

ϕ = relative froth density, ratio of froth density to clear liquid density

(13-49)

(∗From Hutchison, et al., Ref., 11 in Ref. 193)

Use β for design pressure drop calculations [193].

where

F_{va} = vapor flow parameter based on active area,

(13-50)

h_1 = equivalent height of clear liquid on tray, in.

h_f = height of froth (aerated mass) on tray, in.

h_{ow} = height of liquid crest over weir, measured from top of weir (straight or circular), or from bottom of notches (v-notch weirs) , in.

h_w = height of weir above tray floor, in.

v_a = vapor velocity based on active area, ft/sec.

β = aeration factor, dimensionless, Figure 13-64

13.42. Total Wet Tray Pressure Drop

A. Conservative:

This will give a higher pressure drop per tray than method (B).

B. Hughmark and O'Connell Method:

The results of this approach agree with a considerable number of tests reported over a wide range of operation.

(13-51)

C. Fair Method from Reference 193 (used by permission)

Total pressure drop across the tray:

For weeping sieve trays, see Figure 13-69 and Figure 13-70 and example in later paragraph.

Figure 13-69

Weeping correlation for sieve trays with downcomers.

Used by permission, Hughmark, G. A. and H. E. O'Connell, The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 53, p. 127M (1957). All rights reserved.

h_h = head loss due to vapor flow through perforations, in. liquid, see Equation 13-45.

Determine C_o from Figure 13-67 [193]. Note that determining froth density is by experiment and/or best judgment based on similar or related systems. Usually $\phi \cong 1.0$ to 0.50. The method confirms literature and industrial tests to about ± 15%.

13.43. Pressure Drop through Downcomer, h_d

Calculate as for bubble cap tray.

13.44. Liquid Backup or Height in Downcomer

(13-52)

Note that if an inlet tray weir is used, the $(h_w + h_{ow})$ group is replaced by the corresponding $(h_w' + h_{ow}')$ calculated for the inlet weir using the

same algebraic relations.

Free Height in Downcomer

(13-53)

Minimum Vapor Velocity: Weep Point

The “weep point” is considered to be the minimum vapor velocity that will provide a stable tray operation, preventing liquid from passing through the holes and bypassing the overflow weir and downcomer.

This point is generally considered the lower point of operation for the tray which maintain acceptable efficiency. Some systems are known to operate at only slight reduction in efficiency with vapor velocities well below the weep point values. It is impossible to predict this behavior at present. Weeping is usually the limiting condition in design for low vapor rate, high liquid rate systems. Some factors affecting the weep point of any system are described in the following sections.

Weeping and dumping are types of drainage that occur during tray operation, and are more sensitive in the operation/control of sieve trays than for valve or bubble cap trays. Lessi [194] presents an analysis of these conditions. Weeping simply means that the gas/vapor volume passing upward through the tray is not sufficient to prevent liquid on the tray from running back down to the tray below, thereby affecting the tray efficiencies. Dumping is a term more associated with sieve trays than the others; however, in concept it represents a large or excessive amount of liquid draining off the tray, greater than weeping, and could be considered a type or forerunner of flooding of the column.

Hsieh and McNulty [210] developed a new correlation for weeping of sieve and valve trays based on experimental research and published data. For sieve trays the estimation of the weeping rate and weep point is recommended using a twophase countercurrent flow limitation model, CCFL.

The procedure [210] for weeping calculation and determination of vapor rate that will result in a certain weeping rate is (used by permission, Reference 210 *Chemical Engr. Progress*, all rights reserved):

a. Calculate Z

(13-54)

b. Use values of m and C as determined by Reference 210

For sieve trays

m = 2.01

C = 0.74

For type T-valve style (Koch)

m = 2.87

C = 0.74

For type A-valve style (Koch)

m = 2.01

C = 0.74

c. Calculate $j*_L$

(13-55)

using the given volumetric *weeping* liquid rate, and total hole area.

d. Calculate $J*_G$

(13-56)

$J*_G = C$

e. Calculate vapor rate, V_G, based on value of in (d) above using:

(13-57)

f. Using V_G from (e) above based on superficial gas rate, calculate $V_{G,\ weep}$ based on total perforated hole area only, ft/sec; that is:

ft^3/sec vapor – (V_G, ft/sec) (A_H, ft^2)

The weep point for sieve or valve trays is the vapor rate at which the liquid weeping rate is diminished to zero. Thus, $J*_L$ approaches zero as $J*_G$ is increased [210]. For a vapor rate that leads to $J*_G$ higher than the weep point value, there should be no weeping.

$W_{index} > 0$, for no weeping

The higher the value of W_{index}, the more confidence that there will be no weeping [210]. At a constant weep point, $J*_G$, the higher the percentage opening of the tray, and the higher will be the vapor volumetric flow required to satisfy the weep point criteria.

To calculate the weep point, use $J*_G = 0.74$ and calculate Z from (a)

above, then calculate V_G from (e) above.

Ludwig's [210] report of test results shows that below the weep point for Type-T and Type-A valve trays, a consistently low weeping rate can be maintained, while for sieve trays the weeping rate increases rapidly at low gas flow. For similar operating conditions, the weeping rate for a valve tray can be an order of magnitude lower than the corresponding weeping rate for a sieve tray with the same open area. The tests assume uniform weeping across the entire tray deck; however, recent tests [210] indicate that for a larger 8-ft diameter (versus 3-ft) tray, weeping occurs preferentially along the periphery of the bubbling area, indicating that for the larger diameter, the actual weeping rate can be lower, by more than 30%, when referenced to the present models prediction.

The equivalent hole diameter for use in the equation for Z when considering the two types of valve trays studied here is given by:

(13-58)

(equivalent hole diameter, in.), see Equation 13-60

For Type-T valves; $R_V = 1.0$

For Type-A valves; $R_V = 0.8$

For the valve tray equivalent hole diameter, see illustration.

(13-59)

(13-60)

(13-61)

(13-62)

(13-63)

Note: all *dimensions* in inches

where

A_H = total hole area, ft^2
A_{op} = open area of tray, ft^2
A_{VM} = maximum valve open area, ft^2
C = empirical constant in CCFL correlation
C_L = liquid phase loading factor defined in Equation 13-66, ft/s
C_{LW} = measured rate of weeping.
C_v = gas phase loading factor defined in Equation 13-65, ft/s
D_H = hole diameter, in.
D_{HE} = equivalent hole diameter, in., Equation 13-63
D_v = valve diameter, in.
F_v = valve tray F-factor, ft^3/min/valve

F_{VM} = valve tray F-factor at the beginning of the valve open region, ft^3/min/valve
g = gravitational constant, ft/s^2
h_c = clear liquid height, in.
h_D = dry tray pressure drop, in.
H_{GAP} = maximum lift of a valve, in.
H_{OP} = valve lift, that is, the distance between the bottom of a valve and the top of the tray deck, in.
J_G* = dimensionless gas velocity
J_L* = dimensionless weeping liquid velocity
m = empirical constant in CCFL correlation
N_v = valve density, number of valves/ft^2
Q_{GV} = volumetric gas flow rate to a valve, ft^3/min/valve
R_v = fractional opening in the circumference of a valve
V_G = superficial gas velocity in channel (not tower), ft/s.
V_{GH} = gas phase superficial hole velocity, ft/s
V_L = superficial liquid velocity in channel (not tower) ft/s∗, for sieve trays, divide total vapor volume by total perforated hole area
V_{LH} = liquid phase superficial hole velocity, ft/s∗, for sieve trays, divide total liquid flow by total perforated hole area
z′ = Laplace capillary constant
Z = characteristic length in CCFL model
∗ for valve trays, see calculation analysis in text
Greek letters
ρ_G = gas density, lbm/ft^3
ρ_L = liquid density, lbm/ft^3
σ = surface tension, dyne/cm

Figure 13-71Figure 13-72Figure 13-73 and Figure 13-74 illustrate the

correlation of the data with the proposed model and resulting design procedure. Additional illustrations accompany the reference. For Figure 13-71, the C_V and C_L parameters are plotted. For sieve trays, the actual hole velocities are used; where for the Type-T valve tray the "hole" velocities are calculated based on the maximum open area, A_{VM}.

(13-64)

(13-65)

(13-66)

(13-67)

Figure 13-70

Weeping sieve trays.

Used by permission, Smith, B. D., Design of Equilibrium Stage Processes, Chapter 15 by J. R. Fair, McGraw-Hill Book Co. (1963), All rights reserved.

Figure 13-71

Weeping performance comparison. (Valve tray also gives a lower weep rate at a liquid flow rate of 50 gal/min/ft of weir).

Used by permission, The American Institute of Chemical Engineers; Hsieh, C-Li, and K. J., McNulty, Chem. Eng. Prog., V. 89, No. p. 71 (1993). All rights reserved.

Figure 13-72

Weeping data for 9.4% open area sieve tray.

Used by permission, The American Institute of Chemical Engineers; Hsieh, C-Li, and K. J., McNulty, Chem. Eng. Prog., V. 89, No. p. 71 (1993). All rights reserved.

Figure 13-73

Weeping rate type – T valve (Koch) vs. sieve tray.

Used by permission, The American Institute of Chemical Engineers; Hsieh, C-Li, and K. J., McNulty, Chem. Eng. Prog., V. 89, No. p. 71 (1993). All rights reserved.

Figure 13-74

Weeping correlation for type – A valve (Koch) tray.

Used by permission, The American Institute of Chemical Engineers; Hsieh, C-Li, and K. J., McNulty, Chem. Eng. Prog., V. 89, No. p. 71 (1993). All rights reserved.

Hsieh and McNulty [210] (also see section on Sieve Trays) show that the weeping rate for a 14.3% open area valve Koch Type-T (Figure 13-10) is nearly an order of magnitude less than for a corresponding sieve tray with the same percentage open area. The typical valve tray tends to throttle the liquid, as well as the vapor flow as these flows change. As an example of tests, Figure 13-71 [210] compares sieve and valve tray weeping at 50 gpm/min/ft weir with the gas and liquid rates based on the total bubbling area of the tray, A_{TR}. Note that the action of the Type-T valve closes down as the gas flow rate drops, but maintains a low weeping rate within its entire weeping region [210]. This also allows the efficiency of the tray to stay relatively constant over the weeping region.

The weeping rate of the sieve tray is strongly influenced by the gas flow rate; that is, the weeping rate will increase as the gas flow rate reduces below the weep point, i.e., where weeping starts. Note the comparison of sieve and valve trays during weeping, Figure 13-73 [210].

Figure 13-74 [210] correlates weeping for the Type-A (Koch) valve trays, discussed earlier. For more details on the estimated design, see reference cited. The correlation developed with sieve trays still is used, and Equations (13-54)(13-55)(13-56) and (13-57) cover valve trays for rate of weeping and weep point. For columns larger than

about 3-ft diameter, the actual weeping rate can be more than 30% lower than the current calculations indicate. This is largely due to a non-uniform weeping along the periphery of the bubbling area of the tray.

13.45. Weep Point (Velocity)

1. Increases as the liquid surface tension decreases

2. Decreases as the hole size decreases

3. Increases as the plate thickness decreases

4. Increases as the percentage free area increases

5. Increases for hole spacing close to $2d_o$ and smaller. Spacing of $3d_o$ and $4d_o$ gives better operation. Only the ⅛-in. holes of Hunt [33] indicate that $2d_o$ spacing may be acceptable if the holes are very small.
6. Decreases with increasing wettability of liquid on plate surface. Kerosene, hexane, carbon tetrachloride, butyl alcohol, glycerine-water mixtures all wet the test plates better than pure water. The critical tray stability data of Hunt et al., [33] is given in Table 13-11 for air–water, and hence the velocities for other systems that wet the tray better than water should be somewhat lower than those tabulated. The data of Zenz [78] are somewhat higher than these tabulated values by 10–60%.

These values are to be used in establishing first estimates of lower limiting vapor velocities. Actual values should be calculated as outlined in the following.

The two approaches to determining the weep point are:

A. Conservative Design

1. Assume a minimum vapor velocity through the holes.
2. Calculate h_{dt}, Equation 13-43.
3. Compare calculated h_{dt} with value of dry tray pressure drop as given:

4. This is based on the correlation of Mayfield [45] where: h_{dt} (weep) = dry tray pressure drop at tray weep point, in. liquid.
5. Set minimum design dry tray pressure drop 30% above the value of h_{dt} (weep).

B. Normal Design [31]

1. Assume a minimum vapor velocity through the holes. Calculate v_{om} $(\rho_v)^{½}$ (minimum).
2. Calculate wet tray pressure drop, determine effective head from Figure 13-68.
3. Read weep point velocity factor, v_{om} $(\rho_v)^{½}$, from Figure 13-69. The assumed value of v_{om} must be greater than the value read from the curve for
4. Minimum design vapor velocity through the holes may be used as calculated, or if additional safety is required increase the value by 20%.

13.46. Entrainment Flooding

The increasing use of sieve trays in distillation and absorption-stripping has allowed the collection of important performance and design information. Flooding is caused by back-up (buildup) in the downcomer and/or entrainment [183, 184]. When the tray downcomers are sized to carry the liquid load and vapor disengagement in the downcomer (bubbles), the entrainment (jet) flooding is more likely to be the controlling mechanism. If the process application generates froth, this will further complicate the flooding condition.

Most studies have used the Souders-Brown [67] droplet settling velocity concept to calculate entrainment flooding. In this mechanism,

flooding develops due to an upward vapor velocity through the crosssection of the net area of the column that is high enough to suspend droplets, and is expressed as the Souders-Brown flooding constant, C_{SB}, [194, 183, 184].

(13-68)

C_{SB} = flooding constant = C-Factor

(13-69)

The entrainment increases as vapor velocity through the column increases to a power of 2 to 5, so as small as a 10% change in vapor rate results in ten-fold change in entrainment [94]. Low pressure applications usually require lower powers, while higher pressure requires higher powers [94]. Entrainment quantities are sensitive to the vapor velocity. Often, low pressure and vacuum applications develop significant entrainment problems, even when operating below the flood point. Medium to high pressure systems are not often affected except when operating at the flood point.

Generally, when spray entrainment changes to more froth on the tray or in the tray vapor space, then entrainment has been found to increase with liquid rate [94].

As tray spacing increases, entrainment reduces, but does increase with the sieve tray hole diameter [183, 184]. Kister [94] discusses the effects of hardware relationships on spray and froth entrainment.

For sieve trays, Kister's [94, 184] final correlation is for the Souders-

Brown flooding coefficient and is essentially independent of pressure. The Kister and Haas correlation [94] [184] is:

(13-70)

Correcting for the froth-spray regime transition [94]:

(13-71)

(13-72)

where [94]

(13-73)

where

C_{SB} = C-factor at flood, ft/sec

v_{flood} = superficial vapor velocity at flood, ft/sec, based on net area, A_N, ft^2

A_f = fractional hole area, A_H/A_B

A_h = hole area, ft^2 (net)

A_B = bubbling area, column crosssection area less total of downcomer areas, downcomer seal areas and areas of any other non-perforated region, ft^2

A_N = net area (column crosssection area less downcomer top area) ft^2

h_{ct} = clear liquid height at transition from froth to spray regime,

in. liquid.
d_H = hole diameter, in.
S = tray spacing, in.
n = a parameter in the spray regime, in.
σ = surface tension, dynes/cm
ρ_L = liquid density, Ib/ft^3
ρ_c or ρ_v = vapor density, lb/ft^3
Q_L = liquid load, gpm/in. of outlet weir length

Kister and Haas [184] recommend using 25 dynes/cm in Equation 13-70 when the actual surface tension is ≥25 dynes/cm. This correlation is reported [94, 184] to give better replication of physical properties, and predicts most sieve and valve tray entrainment flood data to ±15 to 20%, respectively.

Kister and Haas [184] analyzed sieve and tray data as earlier described [94] and then related their results to application for sieve and valve trays.

Recommended Range of Application: The Kister and Haas [184] Flood Correlation (used by permission, Kister; H. Z. , Distillation Design, McGraw-Hill, Inc., 1992) is

Flooding mechanism: Entrainment (jet) flood only

Tray types: Sieve or valve trays only

Pressure: 1.5–500 psia (Note 1)

Gas velocity: 1.5–13 ft/s

Liquid load: 0.5–12 gpm/in. of outlet weir (Notes 2, 3, 5)

Gas density: 0.03–10 lb/ft^3 (Note 1)

Liquid density: 20–75 lb/ft^3

Surface tension: 5–80 dyne/cm

Liquid viscosity: 0.05–2.0 cP

Tray spacing: 14–36 in. (Notes 4, 5)

Hole diameter: ⅛–1 in.

Fractional hole area: 0.06–0.20 (Note 5)

Weir height: 0–3 in.

Notes:

1. At pressures above 150 psia, downcomer flooding is often the capacity limitation, and is not predicted by the correlation. Caution is required.
2. At high liquid loads (above 7–10 gpm/in.), downcomer flooding is often the capacity limitation, again not predicted by the correlation. Caution is required.
3. Equation 13-73 does not apply for liquid loads lower than 0.5 gpm/in. of weir (35) *. For this reason, this correlation must not be extended to lower liquid rates.

*References in ( ) are from the original source.

4. At lower tray spacing, entrainment flooding may be related to lifting of the froth envelope and to froth rather than spray height. This correlation must not be extended to lower tray spacing.
5. The correlation does not apply when the following three conditions occur simultaneously: (a) Ratio of flow path length to tray spacing is high (>3); (b) liquid rate is high (>6 gpm/in. of weir); and (c) fractional hole area is high (>11%). Under these

conditions, entrainment flooding is related to vapor channeling and vapor cross flow rather than spray height.

Fair's [183] design procedure for establishing an entrainment flooding condition or "point" is as follows: Design Procedure (From Fair, Reference 183, by permission)

The design method presented in this article is best summarized by a stepwise procedure:

1. Establish liquid and vapor flow rates and densities. Obtain or estimate liquid surface tension. If conditions vary significantly across the tower, apply this method to each section of interest wherein conditions can be considered constant.
2. Calculate the flow parameter:

3. Estimate the flood point from Figure 13-75, which accounts for liquid flow effects and is a ratio of liquid/vapor kinetic effects [79]. Flooding velocity is obtained from:

(13-67)

Figure 13-75

Flooding capacity, sieve trays; weir height is less than 15% of tray spacing; low-to nonfoaming system; hole area at least 10% hole sizes -in. to ¼-in dia.; surface tension = 20 dynes/cm.

Used by permission, Fair, J.R., Petro/Chem. Engineer, p. 46, Sept. (1961), reproduced courtesy of Petroleum Engineer International, Dallas, Texas.

Note that the values taken directly from Figure 13-75 apply to sieve trays having a hole area of 10% or more of active area, holes no larger than ¼ in., and liquid surface tension of 20 dynes cm. Corrections are as follows:

a. Hole area

Hole/Active Area	C_{SB} design/C_{SB} chart
0.10	1.00
0.08	0.90
0.06	0.80

b. Hole size

No correction for hole diameter <¼ in. Correction factors for larger diameters not known.

c. Surface tension

4. Choose a tower diameter that will give the desired approach to flooding. Or, if dealing with an existing tower, calculate the approach to flooding.

5. Estimate the fractional entrainment from Figure 13-76 or 13-77.

Figure 13-76

Fractional entrainment, sieve trays.

Used by permission, Fair, J.R., Petro/Chem. Engineer, p. 45, Sept. (1961), reproduced courtesy of Petroleum Engineer International, Dallas, Texas.

Figure 13-77

Entrainment comparison: sieve trays vs. bubble caps for 24-in. tray spacing. Note: BCT = bubble cap tray; ST = sieve tray; FP = flow parameter.

Used by permission, Fair, J. R., Petro/Chem. Engineer, p. 45, Sept. (1961), reproduced courtesy of Petroleum Engineer International, Dallas, Texas.

Using Figure 13-76

(13-74)

Use with Figure 13-75 to estimate both flood point and entrainment.

(13-75)

where

C_{SB} = Souders-Brown capacity coefficient, ft/sec

e = entrainment rate, lb mols/sec

= entrainment rate, lb/lb dry vapor

E_D = dry tray efficiency, fractional

E_W = wet tray efficiency, fractional

FP = flow parameter, dimensionless

G = vapor mass rate, lb/sec or lb/hr

h_f = froth height, in.

h_L = clear liquid height, in.

L = liquid mass rate, lb/sec or lb/hr

L_{MD} = liquid molar rate (without entrainment) lb mols/sec

T = tray spacing, in.

U_N = vapor linear velocity based on area for de-entrainment (usually tower crosssection minus one downcomer), ft/sec

ρ_L = Liquid density, lb/ft^3

ρ_v = Vapor density, lb/ft^3

σ = Surface tension, dynes per cm

σ_w = Surface tension of water, dynes/cm

ϕ =

Ψ = Entrainment expressed as fraction of gross downflow

Fair [183] relates sieve tray valve tray remarks to the extensive work done for bubble cap trays. Figures 13-75 and 13-77 show flooding data for a 24-in. spacing of bubble cap trays from [81] and represents data well for 36-in. diameter columns, and is conservative for smaller columns. Fair's work has been corrected to 20 dynes/cm surface tension by:

(13-76)

For similar systems, where $\phi_1 = \phi_2$, because the flow parameter is related to C_{SB} and ϕ, also, it is affected by:

(13-77)

(13-78)

(13-79)

Figure 13-76 [183] is useful for data correlation, but is not necessary for design purposes [183]. It shows that at high values of parameter FP, sieve trays can be operated very close to the flood point without significant entrainment. Actually, bubble cap trays show the same characteristics [183]. In the low flow parameter region, such trays have a definite advantage, see Figure 13-77. For sieve tray flooding, see data in Figures 13-78 A, B, C, D, E. For reference, the bubble cap entrainment for the 24-in.spacing of trays is in Figure 13-77. Figure 13-78 E is for 18 in.–24 in. tray spacings. All sieve tray charts represent holes <¼ in. and approximately 10% hole area referred to plate area between weirs, or active area [183].

Figure 13-78

A. Bubble cap tray flooding; 24-in. tray spacing. B. Sieve tray flooding, 6-in. tray spacing. C. Sieve tray flooding, 9-in. tray spacing. D.Sieve tray flooding, 12-in. tray spacing. E. Sieve tray flooding, 18-in, -24-in. tray spacing. Studies of sieve tray and bubble cap tray flooding (24-in. tray spacing). (Note that the references listed on the illustrations in Figure 13-78 are from the original source, while ref. 185 is from the text).

Used by permission, Fair, J.R., Petro/Chem. Engineer, p. 45, Sept. (1961), reproduced courtesy of Petroleum Engineer International, Dallas, Texas.

When the hole area is much less than 10% of the active tray area, the flooding limit should be reduced. Fair [183] suggests factors to reduce the $C_{SB}/C_{SB,chart}$. See paragraph 3(a) above.

Figure 13-75 is used for estimating the entrainment-flood point. Liquid particle entrainment is generally considered as reducing tray efficiency (performance).

Figure 13-76 [183] represents the final entrainment correlation used for estimating, thus, based on published data:

(13-80)

The calculated entrainment values may be as good or better than measured values [183]. Figure 13-77 illustrates comparison of entrainment between bubble cap and sieve trays. Fair [183] concludes that for vacuum to moderate pressure applications, sieve trays are advantageous from an entrainment-flooding standpoint.

Example: 13-2: Sieve Tray Splitter Design for Entrainment Flooding Using Fair's Method; (used by permission [183])

For a sieve tray xylene splitter, the following flow conditions are specified:

Liquid rate	200,000 lb/hr
Vapor rate	220,000 lb/hr
Liquid density	46.8 lb/ft^3
Vapor density	0.266 lb/ft^3
Surface tension	16 dynes/cm

From a consideration of contacting requirements, a tower 9.5 ft in diameter is selected. Other pertinent details are: 24-in. tray spacing, 1-in. weir height, -in. dia. holes, 10% hole area (referred to active area) and 8.3 ft^2 downcomer area.

The available vapor flow area is 70.8 − 8.3 = 62.5 ft.^{2} Hence,

From Figure 13-75, C_{SB} for flooding is 0.340. This value is for 20 dynes/cm, 10% hole area and small holes. The only correction required is for surface tension:

From Figure 13-76, $\Psi = 0.055$

If the dry efficiency at this point in the column is 90%, the wet efficiency is calculated by means of Equation 13-75

Experimental flooding and entrainment data for sieve trays are not plentiful, and measurements are not precise. Accordingly, it has been necessary to relate correlations of flooding and entrainment to those of the well-known device, the bubble-cap tray. It appears that the two devices have about the same flooding limits, so long as usual design practice is followed. However, the sieve tray shows entrainment advantages, especially when used in vacuum and atmospheric service.

The flooding capacity for sieve trays has been set into a mathematical equation by Ward [187] using Fair's equation [183] and Figure 13-75. This is turn allows for determination of the column diameter, assuming allowance is made in the flooding velocity to prevent flooding, but perhaps 25% below. Ward [187] compares his curves with Fair's and provides an equation for sieve tray flooding capacity factor:

(13-81)

(13-82)

where

S_{ts} = tray spacing, ft

= liquid mass flow, lb/sec

= vapor mass flow, lb/sec

ρ_v = vapor density, lb/ft^3 at flowing conditions

ρ_L = liquid density, lb/ft^3

Ward [187] reports best fits for curves at tray (or plate) spacing in the range of 0.50 to 3.0 feet, and at the ends of the curves.

By analogy to Fair's [183] work,

(13-83)

Calculate column diameter using U_{flood} reduced by 15–25%, or increase the calculated column area by about 25% and convert to a working diameter.

13.47. Maximum Hole Velocity: Flooding

The maximum hole velocity will give a liquid build up in the downcomer of 50% of the tray spacing.

To determine the maximum velocity:

1. Assume a hole velocity.
2. Calculate liquid height in the downcomer, H_d by Equation 13-52.
3. If H_d = ½ S_t, the assumed hole velocity is satisfactory, if not, repeat until a close balance is obtained.

13.48. Design Hole Velocity

The design velocity for hole selection also sets the minimum tower diameter. To maximize operational flexibility throughout the expected operating range, the following points should be considered:

A. Select a design velocity near the weep point if:

1. The design vapor rate is, or is very close to, the minimum rate.
2. All change in capacity is to be as an increase over design rate.
3. Reduction in efficiency can be tolerated if vapor rate falls to weep point minimum or below.
4. Low tray pressure drop is required, as for vacuum systems. Design with extra caution under vacuum, since data correlations have not been checked in this region.

B. Select a design velocity near the maximum velocity if:

1. The design vapor rate is the maximum expected. Any change will be in the direction of lower rates.
2. High efficiency is required.
3. High pressure drops are acceptable.

13.49. Tray Stability

Figure 13-79 A of Huang and Hodson [30] and Figure 13-79 B can be prepared from an evaluation of limits of tray performance using the relations set forth herein, or as presented in the original reference using a slightly different analysis.

Figure 13-79

A. Typical performance chart; perforated tray with downcomer. Used by permission, Huang, Chen-Jung and J. R. Hudson, Pet. Refiner, V. 37, p. 104 (1958), Gulf Publishing Co., All rights reserved. B. Effects of vapor and liquid loadings on sieve tray performance,
Used by permission, King, C.J., Separation Processes, McGraw-Hill Book Co., Inc. (1971). All rights reserved.

Unstable liquid oscillations on a tray have received only limited examination when compared to perhaps tray weeping, flooding and froth buildup. Biddulph [87] proposes a dimensionless number to predict when twophase liquid–gas oscillations will occur on distillation trays, this predicts the onset of oscillation:

(13-84)

where

B_s = dimensionless group identifier
U = superficial vapor velocity, m/sec
ε = eddy kinematic viscosity, m^2/sec
h_f = froth height, m
ρ_g = gas velocity, kg/m^3
g = gravitational constant, m/sec^2
d = column diameter, m
ρ_L = liquid density, kg/m^3
= relative froth density = h_L/h_f
h_L = clear liquid head, m

The interpretation of criterion for the use of B_s is that:

1. Full-wave oscillations will not occur for values below $B_s = 0.5 \times 10^{-5}$

2. Half-wave oscillations will not occur for values below $B_s = 2.5 \times 10^{-5}$

To counter oscillation effects, Biddulph [87] recommends the use of two vertical baffles made of expanded metal, with approximately 1-cm openings (0.394 in.), installed parallel to the flow path from the inlet weir to the outlet weir, and located at the ⅓ and ⅔ dimensions across the tray diameter. This oscillation phenomenon appears as the vapor rate increases and then the generally "even" layer of liquid excibits violent lateral movements at right angles to the liquid flow. The two primary forms show a peak of liquid at one wall and a trough at the opposite wall (called half-wave oscillation). This condition then reverses.

With increasing vapor rate, the oscillations become more violent, and liquid entrainment increases up to 70%, decreasing the tray efficiency. On sieve trays, extra weeping occurs, up to 150% compared to a stable tray. Full-wave oscillation is represented by a peak wave forming along the center of the tray with a trough at each wall. This position then reverses itself, and is called "full-wave" oscillation. The full-wave occurs at lower vapor rates than half-wave oscillation. Increases in entrainment and weeping also occur, and are most likely to be characteristic of medium-to small-sized columns, particularly those operating at reduced pressure.

To determine the likely possibility of oscillations occurring in a new or an existing column, or even sections of a column, the original article is recommended.

13.50. Vapor Cross-Flow Channeling on Sieve Trays

Kister *et al.* [213] have examined reported cases of cross-flow channeling related to poor sieve tray column performance, and report

that under specific conditions cross-flow channeling does occur. See Figure 13-80 [213] for a diagram of the postulated vapor flow across a tray. It is known to occur for valve trays and bubble cap trays. This condition has not much been studied the open literature; however, several investigators including Ludwig have observed practice in industrial the selective weeping of sieve trays. Referring to Figure 13-80 [213] for sieve and valve trays at low pressure below 72.5 psig (5 barg) when all three listed conditions occur simultaneously [213]:

1. A fractional hole area (or valve slot) greater than 11% of the bubbling area.
2. A ratio of liquid flow path length to tray spacing greater than 2–2.5:1.
3. A liquid flow rate exceeding 50–60 m^3/hr-m of outlet weir (537.9–645.5 ft^3/hr-ft of outlet weir length).

Figure 13-80

Vapor cross-flow channeling. Note entrainment near tray middle and outlet, and weep near tray inlet.

Used by permission, Kister, H. Z., Larson, K. F. and P. E. Madsen, The American Institute of Chemical Engineers, Chem. Eng. Prog. V. 88, No. 11, p. 86, (1992), All rights reserved.

Vapor cross-flow channeling:

1. Intensifies as the fractional hole area increases.
2. Intensifies as outlet weir height increases and as the liquid flow rate increases.
3. For valve trays the effects are observed only for the venturi (low dry pressure drop) valve.

4. For bubble cap trays, the phenomenon is believed to be induced by excessive hydraulic gradient; it is recommended to keep hydraulic gradient to less than 40% of the dry pressure drop.
5. For sieve and valve trays, vapor cross-flow channeling is believed to occur when dry pressure drop is low (low vapor velocities, high fractional hole area and smooth openings) and with a significant hydraulic gradient (i.e., long flow path, high liquid velocities) [213].
6. Is believed to be a froth regime (liquid in continuous phase above the tray and gas present as bubbles in the liquid) phenomenon rather than a spray regime (gas in continuous phase above the tray and liquid present as drops dispersed in the gas), which is usually in vacuum and low liquid-rate service [213].

Large fractional hole area, long flow path relative to tray spacing and high liquid flow rate are the key factors leading to the formation or intensification of vapor cross-flow channeling on sieve and valve trays.

13.51. Tray Layout

Some of the details of tray layout are given in Figure 13-5A. The working details can be set by the required performance.

1. A tower diameter is selected based on Souders-Brown (usually 20−50 percent conservative) or Hunt's relation, Equation 13-33.
2. Assume tray layout as: downcomer areas, non-perforated area; perforated area. Base downcomer requirements on bubble cap tray information of Figure 13-38.
3. Determine the percent hole area in the active tray portion for pressure drop calculation. Note that hole size does not have to be set at this point. (Figure 13-81).

Figure 13-81

Per cent open hole area for perforated and sieve trays.

4. Calculate the expected tray performance.
5. From the selected design hole velocity and the corresponding total vapor rate, the total number of holes can be determined for a given assumed hole diameter.

(13-85)

From Figure 13-82, or by calculation, determine the plate area required for the holes on the pitch selected. Several selections may be tried to be used with the tray layout. These should be checked to agree with the assumed percent hole area of Step 3.

Figure 13-82

Number of holes in perforated plates.

6. If the tray does not balance area-wise, assume a new area arrangement or even diameter, if indicated, and recheck the procedure.

Example 13-3: Sieve Tray Design (Perforated) with Downcomer

The conditions for tray design in a chlorinated hydrocarbon finishing tower are:

1. Clean service, no fouling or suspended material.
2.

	Top	**Bottom**
Vapor rate, ft^3/sec	5.23	5.58
Liquid rate, gpm	9.57	22.1
Vapor density, lb/ft^3	0.582	0.674
Liquid density, lb/ft^3	83	85
Surface tension, dynes/cm	20	20+

3. Tray spacing is to be close as possible, because vertical installation space is a premium.

Estimated Tower Diameter

Souders-Brown method:

(13-1)

From Figure 13-20, C = 100 for 9-in. tray spacing.

In this case, rates are close and ρ_v does not change much from bottom to top of tower:

Tower crosssection area = 5.23/0.36 = 14.5 ft^2

Using the Hunt equation:

Assume: $h_w + h_{ow} = 1.5$ in.

At surface tension = 20 dynes/cm.

For $e_w = 5\% = 0.05$

Figure 13-59 reads: allowable tower velocity = 2.2 ft/sec

Required tower area = 5.58/2.2 = 2.54 ft^2 (bottom, largest)

Select: tower diameter = 2.5 ft

A 2 ft tower would be expected to perform satisfactorily with properly designed trays. However, a 2.5 ft tower is the minimum diameter suitable for internal inspection and maintenance. The cost of a tray tower of 2.5 ft has been found to be no more, and from some bids 5 percent less, than the smaller 2 ft. tower. A 2 ft. tower would either be used with packing or with trays inserted from the top on rods with spacers. This would allow removal of the trays for inspection and maintenance.

Tray Layout Based on 2.5 ft Diameter Tower

Use a segmental downcomer on a cross-flow tray.

From the residence time in downcomers for bubble cap trays, and at the very low tray spacing of 9 inches, select an allowable liquid velocity of 0.1 ft/ sec.

Percent of tower area = 0.492 (100)/4.91 = 10.01%

Using Figure 13-38 for segmental downcomers, at 10% downcomer area, the weir length is 72.8% of tower diameter.

Weir length = 72.8 (30)/100 = 21.8 in.

Since standard details for fabrication are already available for a tray with a 19.5 in. weir in a 30 in. tower (65% of dia.), try this as the first tray to be examined. This is 6.8% of tower cross sectional area. Downcomer area = 0.068 (4.91) = 0.334 ft^2.

Hole Size

Try -in. dia. on ½-in. pitch

This is spacing of 2.66 d_o, and is as close as good design would suggest. Use ⅛-in. tray thickness.

Ratio d_o/c = /½ = ⅜ = 0.375

Percent hole area = 12.8% (of perforation area only) as shown in Figure 13-81.

Minimum Hole Velocity: Weeping

Assume: Submergence = 1.5-in. = h_{s1} = h_{d1} (neglecting Δ/2)

Dry Tray Pressure drop, h_{dt}

The hole diam./tray thickness ratio = /⅛ = 1.5

From Figure 13-66, orifice coefficient, C_o = 0.78, β = 0.128

Effective head

For h_{s1} = 1.5, F_s < 14

Read Figure 13-68, effective head = 1.58 in. Liquid.

Total Wet Tray Pressure Drop

h_t = 0.608 + 1.58 = 2.188 in. liquid

Weep point

Using Figure 13-69, curve A, and h_t = 2.19 in. liquid

Read weep point velocity = 12.5 = v_{om} $(\rho_v)^{½}$

Curve A is used when in doubt, and it gives a higher minimum v_{om}, which is on safer side for design.

Because $v_{om} (\rho_v)^{½} = 12.5$ is less than the assumed value of 13, the 13 will be used.

Maximum Hole Velocity at Flood Conditions

Assume $F_s = v_o (\rho_v)^{½} = 20$ max.

Dry tray pressure drop

= 1.42 in. liquid

Effective head, h_e

= 1.42 in. liquid, for $F_s > 14$ and $h_{sl} = 1.5$

Total wet tray pressure drop, $h_t = 1.42 + 1.4 = 2.82$ inches liquid.

Liquid Back-up or Height in Downcomer

$H_d = 2.82 + 1.5 + 0 + 0$ (assumed, to be confirmed)

$H_d = 4.32$ in. liquid

The limit on H_d for flooding is $S_t/2 = 9/2 = 4.5$ in.

Therefore F_s = 20 appears to be close to minimum.

Design Hole Velocity

Select a velocity represented by F_s factor between minimum and maximum limits.

20 > Design > 13

Select a median value of F_s = 17, because freedom to operate above and below the design value is preferred in this case.

Design Basis

F_s = 17

1. Weir height selected = h_w = 1.0 inch
2. Height of liquid over weir, h_{ow} = 0.52 in.

From Figure 13-42 at 22.1 gpm and l_w = 1.62 ft

3. Submergence, h_{sl} = (f) (h_w) + h_{ow} = (1) (1.0) + 0.52 = 1.52 in. liquid
4. Downcomer pressure loss. Clearance between bottom of downcomer and plate = 1 in. max. Underflow area = (9.5 in.) (1 in.)/144 = 0.065 ft^2. Because this is less than the down flow area (of 0.334 ft^2), it must be used for pressure drop determination. No inlet weir used on this design.

5. Dry tray pressure drop

6. Effective head

h_{sl} = 1.52 in.

h_e = 1.3 in. liquid for F_s > 13, Figure 13-68

7. Total wet tray pressure drop

$h_t = 1.03 + 1.4 = 2.43$ in. liquid

8. Total tower pressure drop for 45 trays

An actual operating tower measured 5 psi ±. It is satisfactory to average the conditions for top and bottom of tower when flows do not vary significantly. Otherwise, parallel determinations must be carried through for top and bottom (and even feed in some cases) conditions.

9. Number of holes required

Hole size selected = -in.

Hole spacing or pitch = ½in.

From Figure 13-82, Holes/in.^{2} plate area = 4.62

Area of a -in. hole = 0.0276 in.^{2}

Calculation Summary

Maximum Velocity	Design Velocity
F_s = 20	17
	= 17 $(0.582)^{½}$
	= 22.3
	= 17 $(0.674)^{½}$
	= 20.7
No. Holes required: CFS at top = 5.23	
CFS at bottom = 5.58	

The selected design F_s = 17 gives the number of holes to operate at

these conditions. Note that the values of 1223 and 1410 holes for the top and bottom respectively indicated operations somewhat closer to the tower maximum than to the weep point. This usually insures as good an efficiency as is obtainable for a given system. It may limit the flexibility of the tower, since there will not be enough holes to operate down to the weep point at the given design flow rates.

On the other hand the tower should be able to operate at changing vapor and liquid loads without serious upset. In this type of tray, the designer has a selection of holes, in this case: For the top select 1,100 to 1,500; for the bottoms, select 1,300 to 1,750, and still expect acceptable performance.

The fabrication of all trays may be punched or drilled (more expensive) with 1410 holes and those in the lower section have blank strips placed over the inlet and outlet edge rows until approximately 1223 holes are left open in the top section above the feed tray.

For close examination of systems having varying latent heats and flow rates, it is wise to examine several points in the tower, even tray by tray in some cases, to be certain that the number of holes in that tray does not place its performance too close to weep or flood conditions. Towers have been built and operated at rated peak loads with every tray having a significantly different number of holes.

10. Mechanical tray layout details. Allow a total of 3½ in. on diameter for extension of tray ring-type support into the tower. This reduces available tray area. Other support details might make more area available. Each must be examined.

Allow 5-in. clearance (no holes) between inlet downcomer and first row of holes. The 5 in. could be reduced to 3 in. minimum if an inlet weir were used.

Allow 3 in. clearance (no holes) between outlet weir and adjacent row of holes.

Downcomer width = 3.6 in. (From Figures 13-38 and 13-83 at 65% weir length).

Figure 13-83

Sieve tray with downcomer layout for Example 13-3.

Area determinations: Figure 13-83.

Area of segment of circle (2) with chord AD:

Diameter circle (2) = 30 – 3.5 = 26.5 in.
Height of chord = 13.25 – (15 – 3.6 – 5) = 6.85 in.
Chord height/circle dia. = 6.85/26.5 = 0.258
Referring to Perry's Handbook, (pg. 32, 3rd Ed.)
Area = 0.161 $(26.5)^2$ = 113.2 $in.^2$

Area of segment of circle (2) width chord BC:

Height of chord = 13.25 – (15 – 3.6 – 3) = 4.85
h/D = 4.85/26.5 = 0.183
Area = 0.0984 $(26.5)^2$ = 69.1 $in.^2$
Area of circle (2) =
Area available for holes = 552 – 113.2 – 69.1 = 369.7 $in.^2$
Area required for holes = (1410)/4.62 holes/$in.^2$ = 305 $in.^2$

Actually not the entire tray needs to be drilled. However, the location of “dead” or imperforated areas must be carefully selected, preferably next to weirs. A special punching (or drilling) arrangement for the holes can run the cost of the trays quite high. It will probably be preferable to check effect of punching holes in entire area A B C D of Figure 13-83.

Area = 369.7

No. holes = 369.7 (4.62) = 1,710 holes

This number is in the range of acceptable performance for bottom section and should be punched. If performance indicates that fewer holes are preferable, blanking strips can be added (or even added before the trays are installed). The top trays would require blanking of holes.

Example 13-4: Tower Diameter Following Fair's Recommendation in Smith [193] (used by permission)

Following the example of Fair [193], the technique is summarized (used by permission, as paraphrase, not copied directly, with a different example):

Design a sieve tray column to separate benzene and toluene to produce 35,000 lb/hr of benzene as overhead product at atmospheric pressure, and use a reflux ratio of 5:1 (reflux returned to net overhead). Using the top tray as the first design basis, which can be followed by other points in the column, determine by the material balance:

Top tray data:

Material: 95%+ benzene

Molecular weight: 78.1

Operating pressure: 14.7 psia
Operating temperature: 176°F
Liquid density: 43.3 lb/ft^3
Vapor density: 0.168 lb/ft^3
Liquid surface tension: 21 dynes/cm
Maximum liquid load: 5(35,000) = 175,000 lb/hr (504 gpm)
Maximum vapor load: 210,000 lb/hr ≅ (347 ft^3/sec)
System: Nonfoaming, noncorrosive, non-fouling

Tower Diameter

Flow Parameter:

(13-81A)

Tray Spacing and Design

Select as quite common; -in. dia. holes, with hole area/tower area = 0.10, 14 US Std. Gauge stainless steel tray material, which is 0.078 in. thick, 2 in. weir height, and 24 in. tray spacing.

Select single cross-flow tray, segmental downcomers, and straight weirs, with weir length equaling 77% of tower diameter. The downcomer segment is 12.4% of the tower area.

Diameter

From Figure 13-75 read,

C_{SB} = 0.36, at F_{lv} at 0.0519 (previous calculation)

The system surface tension is approximately 20 dynes/ cm, therefore no correction is necessary.

In this system, use 85% of flooding condition for design:

(13-67)

At 85% flood:

ft/sec, vapor velocity based on net crosssectional area for vapor flow above tray, usually, $(A_t - A_d)$, ft^2

V = 347 cfs

At 88% total tower area, $A_d = 0.12\ (A_t)$, ft^2; downcomer area

Tower diameter, for 85% flooding.

For fabrication convenience and practicality, select, D = 10.5 ft

Actual A_t at 10.5 ft = 86.59 ft^2

Mechanical Features

A_t	= 86.59 ft^2
A_d	= 0.12 (86.59) = 10.39 ft^2

A_n	= Net crosssection area for vapor flow above tray
$(A_t - A_d)$	= 86.59 – 10.39 = 76.2 ft^2 (usual situation)
A_a	= active or bubble area of tray
$(A_t - 2A_d)$	= [86.59 – 2 (10.5)] = 65.59 ft^2
A_h	= net perforated area of tray, ft^2 = (0.10) (86.59) = 8.66 ft^2

Actual Flow Conditions

Vapor velocity based on net area = 347 $ft^3/s/A_n$

U_n = 347/76.2 = 4.55 ft/sec.

Approach to flooding = $[U_{n,design}/U_{n,flood}]$ [100]

= [4.55/4.90] (85) = 78.99%

Entrainment

Refer to Figure 13-76, Fractional Entrainment, Sieve Trays [183] and for F_{lv} = FP = 0.0519, and 77.4% of flooding.

ε = Liquid entrainment, lb mols/hr
ε = [0.06/(1 – 0.06)] (175,000/78.1) = 143 mols/hr
= 11,168 lb/hr

Pressure Drop

With a hole/active area ratio = 8.66/65.59 = 0.132

With a tray thickness/hole diameter ratio = 0.078/ (3/16) = 0.416

Orifice coefficient, Figure 13-67, read at 0.41 tray/hole gives C_o orifice coefficient = 0.75

Hole velocity = 347/8.66 = 40.06 fps

Dry Tray pressure drop

$h_h = 0.186\ \rho_v/\rho_1\ (v_o/C_o)^2$

$= 0.186\ (0.168/43.3)\ [40.06/0.75]^2 = 2.06$ in. liquid

Weir flow = 0.77 (10.5) (12 in/ft) = 97.02 in.

Weir length = 8.085 ft

Weir crest @ 504 gpm, L_g (see Figure 13-42):

In such a large column, the weir constriction factor (Figure 13-43) is not significant and is not applied to the above h_{ow}.

Aeration:

From Figure 13-64:

for active area

Read figure; aeration factor,

Then, the wet-tray pressure drop is:

1. Operating liquid seal loss, clear liquid on tray

2. Total tray pressure drop

Weep Point

Surface Tension Head:

Then: $A_h/A_a = 8.82/65.59 = 0.134$

Referring to Figure 13-70:

, in. liquid, height of clear liquid at overflow weir

h_{lo} = 2 in. + 1.45 in. = 3.45 in.

$h_h + h_\sigma = 1.98 + 0.103 = 2.08$ in. liquid

Reading the intersection of 3.45 vs. 2.08 shows that for either weep point curve, the weep point is well below the values for operation, so this design is not near the weep point.

Turndown Ratio

By trial and error the tray can be examined to determine the rates that will coincide with the weep point. Thus, the entrainment can establish

the upper limit of operation, and the liquid weeping through the perforations represents the lower limit of stable operations; that is, turndown is generally used to represent the ratio of the maximum allowable liquid rate (at flooding) to the minimum allowable operating throughput.

Downcomer liquid handling:

Based on clear liquid, downcomer velocity:

Referring to Table 13-10 for low foaming hydrocarbons on 24-in. tray spacing, this velocity of 0.106 ft./sec. is quite “safe” compared to a suggested range of 0.55-0.60 ft/sec.

Based on tray spacing of 24 in., assume 50% downcomer full, then:

height of liquid = 12 in. = 1 ft-0 in. then

residence time = 1 ft/0.106 ft/sec. = 9.43 sec

This is compared to about 3 sec reported by Bolles [190]. This should be checked, and the tray spacing may have to be increased, depending on the recalculation for the entire tray.

Liquid Gradient

Referring to equations for aerated liquid pressure drop,

Velocity of froth:

Note Tower diameter = 10.5 ft

Average length for l_{fw} = 18.58/2 = 9.29 ft

Reynold's Modulus:

From Figure 13-65, read friction factor:

f = 0.018 approximate extrapolation

Area of downcomer flow segment:

(13-86)

From Figure 13-38:

At 77% weir times tower diameter, then downcomer area = 12.4% of

tower area, or 18% of tower diameter is downcomer width (depth, *i.e.* weir to wall = 0.18 (10.5) = 1.89 ft for one downcomer). Then, net free area between weirs = 10.5 – 1.89 – 1.89 = 6.72 ft

This is low and should not be a problem across the tray.

Downcomer backup: Assume 1½ in. clearance between bottom edge of downcomer and tray floor (or equivalent depending on the design of downcomer tray relationship). See Figure 13-1.

(13-87)

Head loss through downcomer underflow:

(13-88)

Downcomer backup: See Equation 13-27;

(13-89)

= 2 + 1.45 + 0.0278 + 0.747 + 3.98 = 8.20 in. liquid backup

This is satisfactory, because it is less than 50% of the tray spacing of 24 in. Therefore, the tray appears to have adequate liquid handling capacity. No hole blanking strips are required.

13.52. Perforated Plates Without Downcomers

Perforated plates without downcomers have only recently been included in commercial equipment. The data for rating the performance is not adequately covered in the literature. Since the present developments in industrial equipment have not been released, the information included here is based only on available data. It may serve as a basis for rating, because the basic nature of the contact is quite analogous to the sieve tray. The limits of performance are not well defined, therefore the methods outlined cannot be considered firm. However, they are adequate for many applications and as the basis for further study.

The action of the perforated tray (Figure 13-84) is one of simultaneous flow of vapor and liquid through different holes on a tray; they do not flow countercurrently and simultaneously through the same holes. For a tray in its operating range, the liquid-vapor bubble mixure is in constant agitation. There is usually a layer of relatively clear liquid on the tray with a bubbling, agitated mass outopofit, part of which becomes frothy and/or foamy in appearance depending upon the tray operation and the fluid system properties. There are wavelets of froth-liquid mixture moving from one place to another over the tray. As the head builds up sufficiently to overcome the tray hole pressure drop, the vapor stops flowing in the region and liquid drips and drains through. As soon as the head is reduced, the draining stops and bubbling starts. This action is taking place randomly over the tray. Sutherland [69] observed that vapor was flowing through 70–90% of the holes, well distributed over the plate.

Liquid flowed through the remaining 10–30% of the holes.

Figure 13-84

Perforated trays without downcomers.

The only available data for correlation is that of Sutherland on air-water [69] and of Myers [47] on two hydrocarbon systems, the latter data being at close tray spacings for laboratory columns.

These trays are somewhat sensitive to rapid changes in tower conditions. Towers containing over 40 trays must be controlled within fine limits.

The perforated plate, punched plate, or Dual-Flow plate are terms used to refer to a tray operating without downcomers, with vapor and liquid passing in countercurrent through perforations in the tray. The Dual-Flow term has been coined by Fractionation Research, Inc., and its design know-how is restricted to contributing members, and cannot be presented in this book.

13.53. Diameter

There is essentially no published work on specific tests with these trays as relates to entrainment, *etc.* However, the very close similarity between a perforated plate without downcomers and one with downcomers is sufficient to justify using some data for one in the design of the second. This is the case with diameter determination. The relation of Equation 13-33 for the perforated tray or sieve tray with downcomers can be used for the plate without downcomers.

Generally, the liquid level and foam/froth height will be higher on this tray, hence the value of h_c, clear liquid on the tray, may range from l in. to 6 in. depending on the service.

13.54. Capacity

In general, the vapor capacity for a given tray diameter is 10–35% greater than bubble cap trays and somewhat greater than sieve trays with downcomers. The flexibility or range is limited because reasonable efficiencies fall-off near the dump point for most systems. Usual designs limit the lower operating point to 60–70 % of the flood point, unless particular data is available to safely allow reduction in lower limits without the accompanying loss in tray efficiency.

13.55. Pressure Drop

The pressure drop of these trays is usually quite low. They can be operated at an effective bubbling condition with acceptable efficiencies and low pressure drops. For more efficient operation, the clear liquid height on the tray appears to be similar to the sieve tray, i.e., 1.5-2 in. minimum. This is peculiar to each system, and some operate at 1 in. with as good an efficiency as when a 2 in. is used. When data is not available, 2 in. is recommended as a median design point.

13.56. Dry Tray Pressure Drop

As should be expected, the relation of Hughmark [31] correlates with the data of Sutherland [69] quite well.

(13-90)

The orifice coefficient can be read from Figure 13-66. Sutherland used C_o = 0.85 and 0.73 for -in. and ¾-in. holes respectively, in ¼-in. plate.

13.57. Effective Head, h_e

Although Sutherland did not obtain an equation for total tray pressure drop, correlation at this time indicates that it follows the effective head concept of Hughmark. This is a limited evaluation because the data available did not indicate any clear liquid heights over about 0.75 in.

When "head of liquid" is considered as "clear liquid on the tray", Figure 13-68 may be used to read the effective head, h_e. Values of h_{cl} beyond 1 in. have not been checked due to lack of data, but do agree generally with the plotted results of Sutherland [69].

13.58. Total Wet Tray Pressure Drop

For the data checked:

(13-91)

These results cannot be expected to correlate for a tray just becoming active (very low liquid on tray, 0.1 in. ±), but have been satisfactory at 0.2 in. for clear liquid height, h_{cl}.

To determine tray operation with respect to pressure drop, the value of h_{cl} must be assumed at a reasonable value; the larger the value the better the contact, and higher the pressure drop. Values of h_{cl} should be limited to about 4 in., following sieve tray practice.

13.59. Hole Size, Spacing, Percent Open Area

Hole size is as important in perforated plates without downcomers as for the sieve tray. Published data limit a full analysis of the relationships; however, the smaller holes, ⅛-in., -in., ¼-in. appear to give slightly higher efficiencies for the same tray spacing [47]. Unfortunately the data [69] for the larger ¾-in. holes were not evaluated for efficiencies. Experience has indicated efficiencies equal to or only slightly, 10–15%, less for ¾-in. holes when compared to -in. holes for some systems. Holes as small as -in., -in. and -in. were considered unsatisfactory for high surface tension materials such as water [47].

Sutherland reports frothy type contact for -in. holes and jetting spray bubble action for ¾-in. holes.

Percent open tray areas of 20–30% appear to be optimum for hydrocarbon systems [47].

The larger holes are recommended for high surface tension liquids.

Holes are usually spaced a minimum of $2d_o$, with $3d_o$ to $4d_o$ being preferable. The distance between holes should never exceed 3 in. Thin plates appear to be preferable to thick.

13.60. Tray Spacing

The height of the liquid/froth mixture on the tray is important in

determining tray spacing, as tray flooding moves up the column as the liquid mixture of one tray approaches the underside of the tray above. Tray spacing is recommended as twice the maximum design height of liquid/froth mixture on the tray h_{al}. Spacing of 9, 12, 15, 18 and 24 in. have been used with good results. The closer the spacing, the less the tray flexibility. 15 in. spacing is usually a good design value.

The height of aerated liquid/froth mixture on the tray, h_{al}, (in.) was determined as in agreement with the following relation [69] for air–water for 23% and 40% open trays.

(13-92)

(13-93)

L = Liquid rate, lb/ (hr) (ft^2 of active plate)

This relation does not hold for plates having 10% open hole area, as the heights are several times the corresponding heights for 23% and 40% trays at the same vapor rate, F_c.

For water, the total height of aerated mixture relative to the height of clear liquid on the tray, h_{al}/h_{cl}, had values between 10 and 3, higher values being obtained from the -in. (smaller) holes. Liquid flow rate does not appear to influence these values to any extent.

Higher open tray areas tend to produce a spray rather than a froth.

High vapor rates produce a spray, while the higher liquid rates produce a froth [69].

If these trays are used in systems with exceedingly high foaming tendencies, tray action may be unaccepted impaired. In such cases, the foaming tendency should be examined experimentally. Antifoam agents have proven quite helpful in some problem cases when using these trays.

13.61. Entrainment

Data are unavailable to distinguish between the entrainment of sieve and perforated trays without downcomers. The relation of Hunt *et al.* [33] given for sieve trays is recommended, and should apply quite well.

Sutherland [69] reports for air-water entrainment of 0.0001 to 0.1 lb liquid/lb vapor, averaging 0.01 for 15 in. tray spacing at hole velocity F_s values of 3 to 15. . These values are 1–10% of bubble cap plates. Simkin, *et al.* [64] report a comparison with the Turbogrid tray giving only 3–60% of the entrainment of bubble caps over a wide range of operation.

Sutherland's [69] relation for air-water on a 15 in. tray spacing, correlating ¾ in. holes on 40% and 23% open area, and in. holes with 23% open area is:

(13-94)

where e_w = entrainment, lb liquid/lb vapor.

The correlation of in.

holes in 40% open trays is

(13-95)

Why this deviates from the previous correlation is not understood.

13.62. Dump Point, Plate Activation Point, or Load Point

These trays will dump liquid excessively through the perforations, giving exceeding low efficiencies [47] unless a minimum vapor rate is maintained for a given liquid capacity. The smaller the holes the lower the dump point (vapor velocity).

Figure 13-85 indicates minimum values of F_h to initiate acceptable bubbling tray action. Efficiency at this activation or load point might be expected to be low; however Myers' results indicate good values at this rate.

Figure 13-85

Vapor and liquid rates for tray activation; perforated trays, no downcomers. Compiled form data from Sutherland [69] and Myers [47].

It is recommended that trays be designed for a minimum of 10% above the lower plate activation values. Below these values the tray will dump liquid and become inoperable.

Efficiency

Tray efficiency is as high as for bubble caps and almost as high as sieve trays; higher than bubble caps in some systems. Performance indicates a close similarity to sieve trays, since the mechanism of bubble formation is almost identical. The real point of concern is that the efficiency falls off quickly as the flow rate of vapor through the holes is reduced to close to the minimum values represented by the dump point, or point of plate initial activation. Efficiency increases as the tray spacing increases for a given throughput.

Myers found only a slight decrease in efficiency with an increase in hole size. Industrial experience indicates that large holes of ½ in. and ¾ in. can be designed to operate as efficiently as a small hole say in.

Efficiency appears to fall off significantly for open tray areas above 30%. The higher efficiencies are usually obtained in the 20–25% range of open hole area [47].

Higher efficiencies are obtained for operating condition within 85–95% of the tray flood point.

Flood Point

At the flood point, liquid continues to flow down the column, but builds up at a greater rate from tray to tray. Sutherland [69] demonstrated that flooding moves up the column from the point of origin. For this reason it is important to design perforated trays without downcomers with extra care, as changing internal rates are quickly reflected in performance if the proper hole requirements are not met. They are a useful tray for steady state operations.

13.63. Tray Designs and Layout

1. Establish a tower design diameter using the Souders-Brown method or the relation of Hunt, both given previously.

2. Determine the vapor and liquid rates in the tower at all possible critical points of change. The anticipated maximum and minimum values must be defined.
3. Determine the values of the plate activation velocities (or load points), , for the minimum as well as maximum liquid loads at top and bottom of the tower, and any intermediate points exhibiting significant change in flow rates. For the partial column area trays of Figure 13-84, the v_c refers to the area of active tray limits. If the minimum rates are more than 20% below the maximum, the smaller hole sizes and open areas should be selected.
4. Select a design hole vapor rate, v_o of 1.25 to 1.5 times the minimum values of the plate activation point, or about 25% below the hole velocity at flood conditions.
5. Check the number of holes required at each maximum rate, to determine if the required holes can be placed in the tower area. Use Figure 13-82 to aid in the determination. If more plate area is required than is available, back-calculate the necessary maximum hole velocity. Check if this is reasonable (say not over twice the minimum). If so, the diameter is still acceptable; or change the hole spacing to allow more or fewer holes to be placed in the given diameter of the tower. Use limiting values previously given on hole spacing.
6. Calculate the total wet tray pressure drop, using an assumed height of clear liquid on the tray of 0.5 in. minimum to 4 in. maximum (1 to 2 in. are usual values).
7. Determine the height of aerated liquid on the tray, h_{al}. If the foaming characteristics of the system are less than air-water, the results will be conservative. For systems tending to greater foam and bubbles than the air-water system, approximate a value of h_{al} by multiplying the calculated value by 2, 3, or known relative relationship.

8. Set tray spacing at twice the selected value of h_{al}.
9. Check entrainment at maximum vapor rate.
10. Physical arrangement: refer to Figure 13-84.

For new towers, the designs will usually develop to utilize the entire tower crosssection. However, for existing towers where perforated trays are being installed to replace bubble caps or packing, the optimum active tray area may not utilize the entire crosssection. If the number of holes required is small compared to available area, it is better to group the holes on $2.5d_o$ to $3.5\ d_o$ than to exceed these limits. Holes separated by more than 3 in. are not considered effective in the tray action that i so necessary for good efficiency. Blanking strips may be used to cover some holes when more than required have been perforated in the tray.

If trays are punched, the sharp hole edge side should face the entering vapor.

Example 13-5: Design of Perforated Trays Without Downcomers

A tower separates a weak ammonia solution. Design trays using perforated plates without downcomers for the following conditions as determined from the column performance calculations:

	Top Tray	**Bottom Tray**
Liquid flow rate, gpm	40.8	17.8
Liquid density, lb/ft^3	38.8	54.2
Liquid surface tension, dynes/cm	<13	59
Vapor flow rate, ft^3/sec	5.22	4.3
Vapor density, lb/ft^3	0.593	0.408

Solution

Estimated Tower Diameter

(13-33)

Allowable velocity: assume

(13-34)

From Figure 13-59 for e_w = 0.05 and assumed 15-in. tray spacing at top, tower velocity v_c = 4 ft/sec.

At bottom, v_c = 6.4 ft/sec

Tower area at 4 ft/sec limiting: = 5.22/4 = 1.30 ft^2

(13-35)

Comparison:

Souders-Brown, Figure 13-21 at top tray conditions, which are limiting.
W = 2000 lbs/hr (ft^2) Max. allowable vapor velocity
Top vapor rate = 5.22 (0.593) (3600) = 11,130 lb/vapor/hr
Required area = 11,130/2000 = 5.5 ft^2
Diameter = 2.64 ft, Say 2 ft, 8 in.

Because it is known that the entrainment from perforated trays is considerably less than for bubble caps, the 2 ft, 8 in. diameter would be very conservative and perhaps excessively large.

Tower diameters in the 1 ft, 6 in. to 2 ft range are not usually economical as tray installations. A packed tower might prove the best

economically. Trays can be installed on a central rod and spacer arrangement, with seals between trays and tower shell. Such an arrangement usually brings the cost of the installation up to that of a 2 ft, 6 in. tower. This is the smallest practical size that a man can crawl through.

For the purpose of this design, assume that a cost study has verified the above remarks, and a 2 ft, 6 in. tower will be used. This means that entrainment will be very low on a 15 in. tray spacing. Therefore, a smaller spacing should be considered. From usual fabrication costs, 12 inch spacing is about the narrowest to consider.

The allowable velocity by Hunt for this spacing, , v_c = 4 (8.25/11.25) = 2.94 ft/sec.

Tower area = $\pi (2.5)^2/4 = 4.9$ ft^2

Actual tower velocity = 5.22/4.9 = 1.06 ft/sec

Therefore 12-in. spacing should be adequate entrainment-wise.

Plate Activation Velocities (Minimum)

Top:

Liquid rate, L = (40.8 gpm/7.48) (38.8) (60) = 12,200 lb/hr (ft^2)

From Figure 13-85, read @ -in. holes, 23% open area.

Bottom:

L = 17.8 (54.2) (60)/7.48 = 7,730 lb/hr (ft^2)

From Figure 13-85 read [ ] @ [ ]-in. holes, 23% open area [ ]

Note that Figure 13-85 indicates the operating liquid minimum range is quite stable in the region of design for these trays. The vapor rate must never fall below the above values, or instability will immediately set in and dumping will result.

Design Hole Velocity

Set at 1.5 times the activation velocity, or 2.34 ft/sec at the top and 1.95 ft/sec at the bottom.

Hole Arrangement

[ ] in. dia. on 60°Δ pitch, spaced on ⅜ in. centers.

This gives 22.6% open area (Figure 13-81), close enough to the 23% selected.

No. holes/in^2 = 8.3

Area required for holes:

Top: 5.22/2.34 = 2.23 ft^2

No. holes = 2.23 (144) (8.3) = 2670

Bottom: 4.3/1.95 = 2.2 ft^2

No. holes = 2.2 (144) (8.3) = 2640

Since these are so close, the same number of holes can be perforated in all plates. These should be drilled (punched) in a 1 ft, 6 in. × 1 ft. 6 in. square area as per Figure 13-83. This will reduce costs of trays slightly, as compared to custom perforating to different number of holes.

The tower area is 4.9 ft^2; therefore the entire tray will not be perforated for the conditions of design.

If there is the possibility of vapor and liquid rates being reduced to 50% of the indicated values, this would place the trays as selected above at the dumping point, or activation point, which is not a good operating condition. In this situation, the number of holes should be reduced in order to maintain a velocity of vapor through the holes greater by at least 15% than the activation velocity.

Wet Tray Pressure Drop

a. Dry Tray Pressure Drop

The top appears to be the region of greatest pressure drop:

C_o = 0.82 for -in. hole in -in. tray (Figure 13-66) h_{dt} = 0.0222 in. liquid

b. Effective Head

Assume clear liquid height on tray = 1 in. (note, 1 in. may be a slightly better value than the 1.5 in. assumed when determining e_w).

From Figure 13-68, effective head, h_e = 1.1 in. liquid at F_s = 2.34 $(0.593)^{½}$ = 1.81.

c. Total wet tray pressure drop

h_t = 0.022 + 1.1 = 1.12 in. liquid

Total Tower Tray Pressure Drop

For 15 trays, the maximum expected drop using the expected tray drop at the top:

= 15 (1.12) = 16.8 in. liquid

A more precise approach requires evaluation of the wet tray drop for the bottom condition. However, since it will give a lower value (by inspection), the higher result is preferred as long as a vacuum tower is not being designed. Here, the careful approach is justified.

Also, it must be remembered that the data used in establishing the design criteria are not accurate to better than ±10–20%.

Height of Aerated Liquid on Trays

At top:

h_{al} = 7.06 in. (evacuated as air–water)

Tray Spacing

From this value of h_{al} it is essential that the trays be spaced no closer than 2 (7.06), say 15 in. If lab tests indicate these mixtures foam more than water-air, the value of 15 in. should be increased in relative proportion.

13.64. Proprietary Valve Trays Design and Selection

For proper design of proprietary valve trays (see previous illustrations), the design operating data, basic thermodynamics, and physical properties of the fluids must be submitted to the appropriate manufacturer's technical department for evaluation and final design recommendation. Often the designer for the operating company has established sufficient computer programs to essentially make equivalent studies to the tray manufacturer. Examining the variations in expected tray(s) performance, several studies must be made, and results plotted to analyze the variables and their effects on performance. Billet [208] has reviewed the progress in the design and performance of valve trays.

Klein [201] has evaluated from the literature including manufacturer's, i.e., Glitsch [202], Koch [203], and Nutter [204], design procedures for their respective valve type tray and has developed an approximation procedure suitable for estimating designs. This procedure can be confirmed by the respective manufacturer for each

application. Note that Klein's [201] references to the manufacturer's design manual are to somewhat earlier versions, but are not anticipated to significantly change the estimation by the design engineer. Klein's design method summary follows (by permission).

Dry Tray Pressure Drop

For an operating tray, the pressure drop profile is shown in Figure 13-86 [201]. The valves are "closed" at low hole vapor velocities, although, due to the design of the valves (see Figures 13-10 and 13-12), the metal tabs keep some styles of valves open sufficiently to allow some vapor and some liquid through, even at low flow rates.

In Figure 13-86 point "A" is where the valves on the tray are still "closed" but are just beginning to open. The pressure drop increases with the velocity from "0" to point "A".

Figure 13-86

Typical operating valve tray pressure drop profile. Valves start to open at A, the closed balance point.

Used by permission, Klien, G. F., Chem. Eng. V. 89, No. 9, p. 81, (1982). All rights reserved.

The vapor hole velocity at "A" is [201]:

(13-96)

where

$v_{pt,A}$ = vapor velocity through holes, closed balance point, ft/sec

T_v = metal thickness of valve, in.

R_{vw} = ratio, valve weight with legs/valve weight without legs, dimensions, Table 13-12.

Table 13-11 Tray Stability with Varying Liquid Head, Air–Water System

Used by permission. The American Institute of Chemical Engineers, A.I.Ch.E. Jour. Hunt, C. D. Á et al., V. I. (1955), p. 441. all rights reserved.

Hole dia., in. spacing	**Calculated critical gas velocity in holes ft./sec**			
	$h_{di} = h_w + h_{ow} + \Delta/2$			
	1.0 in.	**1.8 in.**	**2.8 in.**	**3.8 in.**
¼ × $4d_o$	5	25	32	35
¼ × $4d_o$	20	30	45	55
¼ × $3d_o$	27	40	55	70
½ × $4d_o$	25	27	27	30
½ × $6d_o$	30	35	40	45

Table 13-12 Coefficients for the Closed and Open Balance Point Equations: (13-96) and (13-97)

Used by permission, *Chem. Eng.* Klein, G., May 4 (1982), p. 81. all rights reserved.

Valve type	**Flat orifice, R_w**	**Venture orifice, R_w**
3 legs	1.23	1.29
4 legs	1.34	1.45
Caged (no legs)	1.00	1.00
Note: Obtained from measurements on valves.		
C_{VW} = 1.3 for flat and venture valves.		

C_{vw} = eddy loss coefficient, dimensionless, Table 13-12

K_c = loss coefficient, valves closed $(sec)^2$ (in.)/ft^2, see Table 13-

13

$v_a = v_{va}$ = vapor velocity through tray active bubbling area, ft/sec

F_{va} = tray factor based on active bubbling area

=

v_m = valve metal density, lb/ft^3, Table 13-14

Table 13-13 Closed and Open Loss Coefficients for Dry Tray Pressure Drop (13-98) and (13-99).

Used by permission, *Chem. Eng.* Klein, G., p. 81, May 3 (1982), All rights reserved.

Orifice type	Tray deck thickness of		
	0.134 in. (10 gage)	**0.104 in. (12 gage)**	**0.074 in. (14 gage)**
	6.154	6.154	6.154
Venturi	3.077	3.077	3.077
	0.821	0.931	1.104
Venturi	0.448	0.448	0.448

Table 13-14 Common Materials Used for Distillation Valves

Metals	Weight, lb/ft^3
Carbon steel	490
Type 304 Stainless Steel	504
Type 316 Stainless Steel	501
Type 310 Stainless Steel	501
Chrome Stainless Steel, 400 series	484
Monel, 400	551
Nickel	555
Aluminum	170

ρ_v = vapor density, lb/ft^3
g = acceleration of gravity, 32 ft/(sec-sec)
v_h = vapor velocity through valves holes, ft/sec
β = tray aeration factor, dimensionless
ΔP = tray pressure drop, in. liquid
ρ_{vm} = valve metal density, lb/ft^3
t_d = tray deck thickness, in.
ϕ = relative froth density, dimensionless

Note: In Table 13-12 for R_{vw}, the flat orifice refers to a rectangular design valve and the venturi refers to a circular style valve.

The pressure drop remains essentially constant as long as the liquid flow on the tray remains steady during the period point A to point B on the diagram (the open balance point) [201]. At point B, all valves are completely open off their seats, but are on the verge of closing and may be oscillating from open to closed. At point B, the vapor velocity through the holes, opened balance point is:

(13-97)

(13-98)

where
K_o = loss coefficient, valves opened, (sec)^2 (in.)/ft^{2}Table 13-13
$v_{pt.B}$ = vapor velocity through holes, open balance point, ft/sec

Values of R_{vw} and C_{vw} are given in Table 13-12 and ρ_{vm} in Table 13-14. The closed and open loss coefficients for the dry tray pressure drop are given in Table 13-13.

Turndown: It is proposed by Klein [201] that the oscillatory motion of the valve accounts for the greater turndown of a valve compared to other tray-valve designs. The turndown can be controlled by the number of "working" or oscillating valves on the tray, or by changing the uniformity of weight of the valves per tray, or the valve design to obtain different velocities through different valves on the same tray.

Beyond point B on the diagram, the pressure drop for the tray increases as the vapor rate increases. Use Equation (13-99) and (13-100) to determine the dry tray pressure drop, ΔP, in. liquid, Bolles [205] or Klein [201]:

(13-99)

(13-100)

where

v_h = vapor velocity through holes, ft/sec

h_h = dry tray pressure drop, in. tray liquid

ρ_l = liquid density, lb/ft^3

ρ_v = vapor density, lb/ft^3

Aerated – Tray Liquid Pressure Drop

Klein [201] has developed the correlation based on published data of others (his citations) as shown in Figure 13-87. Because the method of

aeration between sieve trays [205] and valve trays is different [201], the same aeration correlation cannot be used; because valve trays have lower liquid ΔP. Figure 13-88 [201] compares the aeration factor for valve, sieve, and bubble cap trays. Figure 13-87 also presents a curve for the relative froth density, ϕ, used for determining froth height as:

(13-101)

(13-102)

(13-103)

Figure 13-87

Correlation for aerated-tray-liquid pressure drop developed from published data for various valves. Note: ϕ = relative froth density. Reference numbers are from original article [201].

Used by permission, Klien, G. F., Chem. Eng. V. 89, No. 9, p. 81, (1982). All rights reserved.

Figure 13-88

Valve trays have the lowest liquid pressure drop of all three types of trays employed (also see Ref. 88, 183, 193 for additional interpretation).

Used by permission, Klien, G. F., Chem. Eng. V. 89, No. 9, p. 81, (1982). All rights reserved.

Hutchinson, cited by Klein [201], developed the relation between β and ϕ:

(13-104)

With this equation, the aeration factor curve β can be developed from the relative froth density curve of Figure 13-87.

Overall tray pressure drop [201]:

(13-105)

(13-106)

where

h_t = total tray pressure drop, in. tray liquid

h_l = aerated tray liquid pressure drop or equivalent clear liquid on a tray, in, tray liquid

h_f = froth height on tray, in.

h_h = dry tray pressure drop, in. tray liquid

h_w = weir height, in.

h_{ow} = crest of liquid over tray weir, in. liquid

β = tray aeration factor, dimensionless

ΔP = tray pressure drop, in. liquid
ϕ = relative froth density, dimensionless
Q_l = liquid flow on tray, gal/min
L_{wi} = weir length, in.
F_{va} = tray F Factor, based on active bubbling area
=
G = vapor rate through all valves, lb/hr

Example: 13-6: Procedure for Calculating Valve Tray Pressure Drop (after Klein [201])

For a venturi type tray, assume the following conditions
Vapor flow, G = 50,000 lb/hr
Liquid flow, Q_l = 205 gpm
ρ_v = 1.91 lb/ft^3
ρ_l = 31.0 lb/ft^3
L_{wi} = 55 in.
h_w = 3 in.
F_{va} = 1.1
Tray froth height: Assume: 12 in.
Per cent of jet flood: 65%
Valve thickness: 16 gage (0.060 in.), 4-legs
Valve material: carbon steel, see Table 13-14.
Valve hole area: 1.65 sq. ft. (separate calculation) = h_o
Tray pressure drop and froth height:

1. Determine $v_{pt,A}$ and $v_{pt,B}$ from Equations (13-95)(13-96) and (13-97)

Close:

$v_{pt.A}$ = 3.06 ft/sec

Open:

2. Determine actual hole velocity, v_h:

Because the actual velocity is operating between the point A and point B, ($v_{pt.A}$ and $v_{pt.B}$):

, for closed valve

= 3.077 (1.91/31.0) $(3.06)^2$ = 2.08 in. liquid

h = 0.448 (1.91/31) $(8.01)^2$ = 1.77 in. liquid. open valve

Because the tray is not near jet flooding, referring to Figure 13-87,

F_{va} = 1.04, then β = 0.61

Overall Tray Pressure Drop:

Froth Height:

h_f = 1.15 in. liquid, (see calculation above)

Using F-Factor, determine β and ϕ from Figure 13-87

at F_{va} = 1.04, then β = 0.61 and ϕ = 0.22

h_l = 2.53 in. liquid (see calculation above)

Calculate froth height, h_f:

in.

Klein [201] refers to Thorngren [206, 207] but suggests that this proposed valve tray flooding is reasonably involved, although considered useful.

13.65. Proprietary Designs

The design engineer cannot adequately design a valve tray that includes operating valves and expects to have reliable performance. The proper approach is to assemble all of the required system/column operating performance requirements and then turn the problem over to a manufacturer who has tested its own valve designs and is capable of predicting reliable performance. The manufacturer can then provide a hydraulic design for the tray, as well as the expected performance of the entire column/tray system. The major manufacturer/designers are Sulzer ChemTech, Nutter Engineering, Harsco Corporation, [204]; Koch Engineering Inc. [203]; Glitsch, Inc. [202], and Norton Chemical Process Products Corporation [233].

There are other manufacturers and engineering companies that are

capable, through good computer programs, of designing competitive distillation designs, and it is not the intent of the above listing to omit any reliable organization, but to simply list the major suppliers in the U.S.

One important point to consider is whether the organization has obtained data on commercially sized equipment, or on that designed and fabricated to their designs, and how the two sets of results compare. The respective design procedure as set forth in each company's design manual will not be outlined in this text, as there is too much detail to produce a reliable tray performance design, and this is included in the manuals. The overall purpose of this section is to allow the designer to (1) become knowledgeable in the component details necessary for a proper design, to be able to communicate with the final designer/manufacturer, and interpret the significance of the final results, and (2) be capable of preparing approximate designs with preliminary information and to develop calculated results to compare with the final designs of others. The designs developed by the methods/procedures presented here are considered reliable for these purposes, and even as final designs, provided there are actual process data and experience to compare then with.

Capps [188] compares valve and sieve tray performance as related to capacity and flooding. Also see sieve tray section presented earlier in this chapter.

Capps [188] examines sieve and valve tray capacity performance and Figure 13-89 [188] is offered for preliminary column sizing, or for determining whether a debottlenecking study is justified. The correlation for flooding, tray rating, and design of a tray is all based on the capacity factor, C_T, equation (Souders and Brown [68] by Capps [188]).

Figure 13-89

Graphical correlation of sieve tray ultimate capacity for hydrocarbons.

Used by permission, Capps, R. W., The American Institute of Chemical Engineers, Chem. Eng. Prog, V. 89, No. 3, p. 35, (1993). All rights reserved.

At total reflux (L/V = 1), Capps found several points in the FRI data that corresponded with this [241] definition of ultimate capacity, i.e., the liquid and vapor load at which any increase in either liquid or vapor would induce flooding by at least one of the following mechanisms:

1. Figure 13-89 shows capacity factor (Souders-Brown velocity) versus system factor (pressure, in this case for hydrocarbons) with L/V as a parameter. In Figure 13-89, the predicted ultimate capacity for a hydrocarbon is obtained by reading the capacity factor at incipient flood for a given pressure at a given reflux ratio, L/V. The Souders-Brown velocity can be used to predict the maximum load achievable for a given column diameter, or, the minimum tower area required for a given load [188, 241].
2. Flood factor is the usual design safety factor (e.g., 80% of flood, $F_{flood} = 0.80$).

(13-107)

(13-108)

(13-109)

Capacity factor, $C_T = V_{load}/A_T$

(13-110)

Using the Souders-Brown factor:

(13-111)

or

(13-112)

Capps analyzes that from Figure 13-89 [188], which was derived from data from Fractionation Research, Inc. in commercial scale tests, a 450-psig de-ethanizer operating at a capacity factor of 0.18 in the rectifying section may not be worth retraying to debottleneck a process, while a 30-psig crude column at a capacity factor of 0.25 may provide a good economic rate of return for a retraying operation/or revamp. These generalized decisions are established by spotting

capacity factors on the chart and noting the potential improvement possible to reach the appropriate L/V curve.

Note that “jet flooding” capacity is fairly insensitive to system physical properties, but that the “system limit” capacity is strongly dependent on physical properties.

Generalized mechanical performance of high pressure and vacuum tray hydrocarbon distillation are shown in Figures 13-90 A and 13-90 B. The representations are for concepts only and do not represent any published data. The charts illustrate the effects of physical properties and pressure on flooding situations. Since flooding is an important condition that limits the performance and capacity of a column, it deserves attention and understanding. The four mechanisms of flooding are [188]:

1. *Jet flooding* occurs due to liquid entrainment induced by vapor jets passing through the liquid flowing on the tray. The entrained droplet may carry into the tray area above and reduce tray efficiency and capacity.
2. *System limit flooding* is similar to jet flooding, due to low surface tension and low density difference between liquid and vapor. Terminal velocity of some entrainment droplets is less than the upward vapor velocity, and hence they are carried up into the tray above, thus reducing tray efficiency and capacity.
3. *Downcomer backup flooding* results from pressure drop at the bottom outlet of the downcomer, causes liquid to back up in the downcomer and flood the tray above. Generally the cause is due to excessive tray pressure drop.
4. *Downcomer twophase flooding* results from vapor failing to disengage from downcomer liquid, and causing twophase flow to pass through the downcomer bottom outlet, causing backup in the downcomer to the tray above. Generally, this occurs in high pressure systems with low surface tension and low density

differences,

where

A_B = bubbling area of tray, ft^2

A_{DCT} = downcomer top area, ft^2

A_T = tower area, ft^2

C_T = capacity factor, based on tower area, ft/s

C_{Tflood} = capacity factor at flood, ft/s

D_T = tower diameter, ft

F_{flood} = flood factor, dimensionless

L/V = internal reflux ratio, dimensionless

m_{vap} = vapor rate, lb/s

$m_{vapflood}$ = vapor rate at flood, lb/s

ρ_{vap} = vapor density, lb/ft^3

ρ_{liq} = liquid density, lb/ft^3

S = tray spacing, in.

V_{load} = vapor load, corrected for density, ft^3/s

Figure 13-90

A. Mechanical performance correlations for high-pressure fractionation trays. Used by permission, Capps, R. W., The American Institute of Chemical Engineers, Chem. Eng. Prog, V. 89, No. 3, p. 35, (1993). All rights reserved. Note: Drawing not to scale. B. Mechanical performance correlations for vacuum fractionation trays.

Used by permission, Capps, R. W., The American Institute of Chemical Engineers, Chem. Eng. Prog, V. 89, No. 3, p. 35, (1993). All rights reserved. Note: Drawing not to scale.

13.66. Baffle Tray Columns

Fair [211] has presented and reviewed many studies of baffle tray, or "splash/shower deck" distillation columns. Figures 13-91 and Figure 13-92 illustrate a simple tray arrangement. The performance of the column is based on the contacting of the up-coming gas/vapor with the liquid cascading from one tray to the one below. The gas must flow through the liquid curtain, and in so doing contacts the liquid for mass and heat transfer.

Figure 13-91

Simple side-to-side baffle arrangement, with liquid flow cascades, for baffle tray column.

Used by permission, Fair, J. R., Hydrocarbon Processing, V. 72, No. 5, p. 75 (1993), Gulf Pub. Co. All rights reserved.

Figure 13-92

Baffle column showing possible enhancements.

Used by permission, Fair, J. R., Hydrocarbon Processing, V. 72, No. 5, p. 75 (1993), Gulf Pub. Co. All rights reserved.

The baffle patterns in the column can be segmental (simple) up to about 4 ft diameter, and larger columns can use a disk and donut

design as in heat exchangers, or double segmented or even multi-segmented as in the layouts discussed under bubble caps earlier in this text.

Pressure drop for 50% cut baffles [211] (see Figures 13-92 and 13-93):

(13-113)

The pressure will be affected by flow rate.

The discharge coefficient, C_v, is often used as 0.6 to 0.7.

These are noted to be high. For 32% and 20% windows (see Figure 13-91 [211]) and curtains, respectively, a coefficient of 0.27 has been determined. The values of C_v from Lemieux's data [212] as presented by Fair [211]:

L, lb/hr-sq^2	**C_v**
0	0.55
3000	0.41
6000	0.30
10,000	0.20
12,000	0.15
15,000	0.15

L, is based on the superficial crosssection of the column, lb/hr-ft^2, and v_w is the linear gas velocity based on the window area, ft/ sec,

(13-114)

The pressure drop data of Lemieux [212] are shown by Fair [211] in Figure 13-93, although there is limited mass transfer data available, Fair [211] has offered this approximate design equation:

(13-115)

Figure 13-93

Pressure drop for 50% cut baffles at 2.0 ft-spacing. the abscissa parameter Z_F is defined as $Z_F = v_W/0.0692\ (\rho_g/\rho_L)^{0.5}$.

Used by permission, Fair, J. R., Hydrocarbon Processing, V. 72, No. 5, p. 75 (1993), Gulf Pub. Co. All rights reserved.

Then, converting to HETP:

(13-116)

where

- a = interfacial area, ft^2/ft^3
- C_l = constant in heat transfer equation = 0.0025 (English units)
- C_g = specific heat, Btu/lb-°F
- C_v = orifice coefficient, Equation 13-113
- G = gas mass velocity, lb/hr-ft^2
- h_g = gas phase heat transfer coefficient, Btu/hr-ft^2-°F

h_L = liquid phase heat transfer coefficient, Btu/hr-ft^2-^{o}F
HETP = height equivalent to a theoretical plate, ft
HTU = height of a transfer unit, ft
L = liquid mass velocity, lb/hr-ft^2
m = exponent $\cong$ 1.0
n = exponent $\cong$ 0.44
Pr = Prandtl number, dimensionless
Sc = Schmidt number, dimensionless
U_s = linear velocity of gas based on total column crosssectional area, ft/sec
v_w = linear velocity of gas based on window area, ft/sec

Subscripts
g = gas
L = liquid
Og = overall (gas cencentraion basis)

Greek letters
ΔP = pressure drop, in. liquid
λ = slope ratio, slope equilibrium line/slope operating line, Equation 13-116
ρ = density, lb/ft^3

Example 13-7: Mass Transfer Efficiency Calculation for Baffle Tray Column (used by permission [211])

Data for example calculation

System	
Mixture	50–50 molar cyclohexane/n-heptane
Total reflux operation	
Operating pressure	24 psia
Temperature	

Temperature	238°F
Relative volatility	1.57
Slope of equilibrium line	1.21
Flow rates	
Vapor F-factor	1.0 ft/sec $(lb/ft^3)^{0.5}$
Gas mass velocity	2,100 lb/hr-ft^2
Liquid mass velocity	2,100 lb/hr-ft^2
Properties	
Liquid density	38.0 lb/ft^3
Liquid viscosity	0.56 lb/ft-hr
Liquid diffusion coefficient	2.40×10^{-4} ft^2/hr
Gas density	0.34 lb/ft^3
Gas viscosity	0.020 lb/ft-hr
Gas diffusion coefficient	0.114 ft^2/hr
Gas Schmidt number	0.516
Gas specific heat	0.294 Btu/lb-°F
Gas thermal conductivity	0.012 Btu/hr-ft-°F
Gas Prandtl number	0.490

For an F-factor of 1.0 ft/s $(lb/ft^3)^{0.5}$, L = G = 2,100 lb/hr-ft^2. For Equation 13-115 a value of C_l is taken as 0.0025. Then, by Equation 13-116 and assuming that most, if not all, of the resistance is in the gas phase:

and

Thus, a 20 ft baffle tray section, with 50% cut baffles on 24 in. spacing can contain 10 elements and produce 5.2 theoretical stages of separation. A corresponding cross flow sieve tray section, with 10 trays at 90% efficiency (16)∗, can produce 9 theoretical stages. This ratio is about as would be expected.

The pressure drop per baffle is:

For the 20 ft section, total $\Delta P = 10 \times 0.11 = 1.10$ in. liquid. The cross flow sieve tray would have a higher pressure drop.

13.67. Tower Specifications

Performance calculations must be interpreted for mechanical construction and for summary review by others concerned with the operation and selection of equipment. Typical specification sheets are given in Figure 13-94A and Figure 13-94B for the tower and internal trays, respectively. Suggested manufacturing tolerances are given in Figure 13-95. A composite cut-a-way view of tower trays assembled is shown in Figure 13-96. A Fractionation Research, Inc. (FRI) suggested distillation tray data sheet is shown in Figures 13-97.

Figure 13-94A Tower specification form.

Figure 13-94B

Tower internals specifications form, tray type columns.

Figure 13-95

Suggested tolerances for distillation type towers.

Figure 13-96

Composite tower-tray assembly illustrating special trays with corresponding nozzles.

Used by permission, Glitsch, Inc.

Figure 13-97

Data specification sheets suggested by Fractionation Research Inc. (FRI) for distillation trays.

Used by permission, Yeoman, N. The American Institute of Chemical Engineers, Chem. Eng. Prog.,

V. 85, No. 10, p. 15 (1989), All rights reserved.

The calculation of nozzle connections has not been demonstrated, but normally follows line sizing practice, or some special velocity limitation, depending upon nozzle purpose.

Tower shells may be ferrous, non-ferrous, stainless alloys or clad (such as monel-clad-steel). The trays are usually light gage metal consistent with the corrosion and erosion problems of the system. The velocity action of vapors flowing through holes and slots accentuates the erosion-corrosion problems, and often a carbon steel tower will use stainless, alloy steel, monel or nickel for trays, caps and all internal parts. Sometimes just the cap or hole portion of the trays are of expensive construction.

When clad metal is used, it is often specified as ⅛ in. or ¼ in. minimum clad thickness. This is usually sufficient to allow proper weld connections. Care must be used in sealing all internal joints of clad material to prevent exposure of the base metal.

The towers are designed in accordance with the particular code (such as ASME Unfired Pressure Vessel) used by the company or required by law. To provide stiffness and bending strength in high winds, design normally factors in a wind load recommended for the area. It is not unusual to design for 75 to 100 mph winds, taking into account the external insulation, piping, ladders and platforms in computing the effective force areas. Foundations must be adequate to carry the total dead weight of the installed tower, platforms, etc., plus the weight of water (or perhaps other fluid) to allow for in-place testing, or complete tower flooding.

Tray types are selected for performance. However, when a particular

type is not specifically required, it is well to consider that, in carbon steel, the trays installed not including tower shell cost approximately:

Bubble Caps, 25% > Sieve, 10% > Perforated without downcomers.

When specifying the mechanical arrangement details, it is important for the designer to take on the role of plant operator in order to analyze what information is needed and how it is to be obtained. For example, once a column is in operation, it is often necessary to determine what may be preventing the system from meeting design specifications for throughput or quality of top, bottom, or side-draw products. To determine such information it is essential to provide at least the following minimum mechanical features to allow extraction of needed samples, temperatures, variations in feed entries, *etc.* (see Figure 13-94A and Figure 13-94B), for example:

1. Thermocouple entry points on about every other tray measure either liquid or vapor temperature.
2. Provide at least three, or perhaps four feed nozzles in addition to the one "theoretically" calculated to be the optimum location. Select these feed locations approximately two and four trays above and below the design basis or theoretical location. These extra nozzles must be oriented on the column so that they have proper feed entry spargers or distributors (entry can be onto the tray or into the downcomer and can be halved from a feed manifold to select the alternate desired location for testing purpose.
3. Reflux nozzles must be arranged to enter the tray with proper designed internal pipe to the tray downcomer or distributor.
4. Pressure tap (couplings) to take several pressure readings in the vapor space above a specific tray up the column. It is better to have too many entries available for testing than too few and not be able to properly examine the column.
5. Sample draw off connections, usually for liquid from the trays,

but some for top (overhead) and reboiler vapor as well as liquid can also be very useful.

Another typical specification with details of the internals of C1007 debutanizer column is shown in Figure 13-98. Figure 13-99a shows the calming section of trays in the column; Figures 13-99b and 13-99c illustrate the layout of CS-HI FI valve trays, and HI-FI CS downcomer of the column respectively. Figure 13-100 shows the process datasheet of the debutanizer column and these figures can also be accessed from Elsevier website.

Figure 13-98

Debutanizer tower specifications.

Figure 13-99A Calming section of trays of Debutanizer column C1007.

Figure 13-99B

Layout of CS-H-FI valve trays of debutanizer column.

Figure 13-99C

Typical HI-FI C-S downcomer of debutanizer column C1007.

Figure 13-100

Process datasheet of debutanizer C1007 column.

13.68. Mechanical Problems in Tray Distillation Columns

Although it appears that a fabricated column with welded internal components, supports, trays, *etc.* should be free from mechanical problems, actual experience proves this is not the case. Most trays are bolted onto supports, and for large columns, tray sections are assembled inside by bolting them together. Actual experience has found that poor column performance can often be attributed to bubble caps and valves knocked (or blown) off position on the trays and often blown to one corner of a tray. Sometimes this condition is found for several trays in a section of the column, thereby preventing any vapor-liquid contact and creating a significant loss of distillation efficiency.

Some of this condition can be attributed to surging or “burping” inside

and pressure surges under increased pressure being created. Other mechanical damage includes nuts coming off bolts, tray metal and welds cracking at or near supports. Corrosion of tray sections and welds is often caused by pressure pulsations at or near the tray's first or second natural frequencies which create vibration and “auto-pulsations”. These occur at resonant or near resonant conditions. Winter [214] presents rough estimating correlations for predicting natural frequencies and deflection of trays.

13.69. Troubleshooting Distillation Columns

To respond to difficulties encountered during operation of distillation columns a very careful and itemized analysis must be made of (a) the process, (b) the mechanical details of the column, and (c) the instrumentation for operation and control. A good column performance designer is generally in an excellent position to examine the operating performance and diagnose the nature and specific location of the conditions that may be preventing or contributing to good column performance. This may often involve detailed computer studies of data for tray-by-tray performance.

It is beyond the scope of this text to thoroughly examine this subject, however, there are several good references including Hasbrouck, *et al.* [215], Kister [117, 347] and Lieberman [349, 350]. This is not an exhaustive list.

Nomenclature

$A = A_T$ = Total crosssectional area of tower diameter, ft^2

A_a = Total annular cap area per tray, ft^2; or

= Active or “bubbling” area of tray, generally $(A_t - 2A_d)$, ft^2, see Figures 13-57 and 13-67

A_B = Bubbling area; column area minus total of downcomer and downcomer seal areas, ft^2 or m^2

A_c = Total cap area inside cross section area per tray, ft^2

A_d = Downcomer area, crosssectional area for total liquid downflow, ft^2; or,
= Minimum flow area at bottom (under) of downcomer per tray, ft^2
A_f = Fractional hole area (actual hole area/bubbling area, A_B)
A_h = Net perforated area of tray, ft^2
A_H = Total hole area, ft^2
A_n = Net open liquid area of one tray, equal to total tower section minus area occupied by caps and risers and minus area of segmental or other downcomer at outlet of tray, ft^2; or = Net area, column area minus area at top of the downcomers, m^2
A_{op} = Open area of tray, ft^2
A_s = Total slot area per tray, ft^2
A_t = Total tower crosssection area, ft^2
A_r = Total riser inside area per tray, ft^2
A'_r = Total reversal area per tray, ft^2
a_a = Annular area per cap, $in.^2$
a_c = Inside crosssection area of cap, $in.^2$
a_d = Crosssection flow area, minimum, of downpipe clearance area between tray floor and downpipe bottom edge, or up-flow area between outer circumference downpipe and any inlet tray weir, $in.^2$
a_o = Individual hole area per hole on sieve tray, $in.^2$
a_r = Riser inside crosssection area per riser, $in.^2$
a_r' = Reversal area per cap assembly, $in.^2$
a_{ro} = Riser outside crosssection area, based on O.D., $in.^2$ per riser
a_s = Slot area per cap, $in.^2$
A_t = Total tower crosssectional area, ft^2
A_{VM} = Maximum valve open area, ft^2
a_X = Smaller area value, a_a or a_r, for use in Equation (13-13) and

(13-15)

B_s = Dimensionless group identifier

C = Factor for Souders-Brown maximum entrainment relation; or

= Empirical constant in CCFL correlation

C_d = Liquid gradient factor

C_g = Specific heat, Btu/lb-°F

C_F = Flooding capacity factor, ft/sec

C_l = Constant in heat transfer Equation 13-115= 0.0025

C_L = Liquid phase loading factor, ft/sec, Equation 13-66

C_o = Orifice (vapor discharge) coefficient for dry tray, Figure 13-66, or 13-67 respectively

C_{SB} = C-Factor at flood (Souders-Brown coefficient), ft/sec (or, m/sec); or

= Souders-Brown flooding constant defined by Equation 13-70

C_T = Capacity factor based on tower area, ft/sec

$C_{T,flood}$ = Capacity factor at flood, ft/sec

C_v = Liquid gradient vapor load correction factor; or

= Discharge coefficient (see accompanying table); or

= Gas phase loading factor, ft/sec, Equation 13-65

C_{vw} = Eddy loss coefficient, dimensionless, Table 13-12

C_w = wet cap pressure drop correction factor, Figure 13-53

c = Hole spacing center to center, in.

D = D_T = Tower inside diameter, ft

D_f = Total flow width across tray, normal to flow, ft

D_H = Hole diameter, in.

D_{HE} = Equivalent hole diameter, in.

D_V = Valve diameter, in.

d = Column diameter, (m)

d_c = inside diameter of cap, in.

d_h = Diameter of weep hole, in. Note that this is the diameter equivalent to area of all weep holes per tray or

= Hole diameter, in. (or mm).
d_o = Hole diameter, in.
d_r = Inside diameter of riser, in.
d_v = Diameter of valve unit at narrowest opening, mm
d_w = Diameter of circular weir, in.
E_d = Dry tray efficiency, fraction
E_W = Wet tray efficiency, fraction
e_w = Weight of liquid entrained per unit weight of vapor flowing, lb/lb
f = Aeration factor (usually = 1.0); or
= Friction factor for froth cross flow, Equation 13-38
f_{hg} = Friction factor for liquid gradient, cross-flow for sieve trays
F = Free height in downcomer above clear liquid level (not froth level)
F_c = Tower velocity factor
F_{flood} = Flood factor, dimensionless
F_h = , for perforated trays, no downcomers
FP = F_{lv} = Flow parameter, dimensionless
F_s = Hole factor = $v_{om} (\rho_v)^{1/2}$
F_{va} = Vapor flow parameter based on active area, defined by $F_{va} = v_a (\rho_v)^{1/2}$, or
= Tray F factor based on active (bubbling) area = (ft/sec) $(lb/ft^3)^{½}$
F_w = Modification factor to weir formula
F_w = Flow parameter, dimensionless
G = V = Vapor or gas flow, lb/hr (see Figures 13-20 or 13-21 or Equations 13-1 or 13-73); or
= Gas mass velocity, lb/hr-ft^2
g = Acceleration of gravity, 32.2 ft/sec-sec

H_{GAP} = Maximum lift of a valve, in.
H_d = Height of clear liquid in downcomer, in.
H_s = Slot height of bubble cap, in.
$(HTU)_{OG}$ = Height of transfer unit, ft
HETP = Height equivalent to a theoretical plate/tray/stage, in. or ft, or possibly mm
h_σ = Head loss due to bubble formation, in. liquid; or
= Head loss due to vapor flow through perforations; in. liquid
h' = Height of overflow weir or bubble cap riser, whichever is smaller, in.
h_{al} = Height of aerated liquid on tray, in.
h_{cw} = Wet cap pressure drop (riser, reversal, annulus, slots), in. liquid
h_c = Head of liquid in bubbling zone; wet cap pressure drop; or taken as in. clear liquid on tray
h'_c = Total dry cap pressure drop, in. liquid
$h_{cl} = h_c$ = Height of clear liquid on tray, in. (or mm)
$h_{cl,i}$ = Clear liquid at the inlet, in.
h_{ct} = Clear liquid height at froth-to-spray transition, in. liquid (or mm)
h_d = Total head loss under downcomer, in. liquid
h'_d = Head loss between segmental downcomer and tray inlet weir, in. liquid
h_{dc} = Head loss of circular downpipe at point of greatest restriction, in. liquid
h_{dcl} = Downcomer height clearance between bottom of downcomer and tray floor, in.
h_{dl} = Dynamic liquid seal on sieve or perforated tray, in. liquid
h_{ds} = Dynamic slot seal, in. liquid
h_{dt} = Pressure drop through dry perforated or sieve tray, in. liquid
h_{du} = Downcomer head loss due to friction and underflow, in.

liquid

h_e = Effective liquid head taking aeration of liquid into account, in. liquid, Figure 13-68

h_f = Height of top of foam above tray floor, in. (or mm)

= Height of free fall of liquid in downcomer; in. or

= Height of froth on tray (aerated mass), in.

h_{fd} = Downcomer backup, in.

h_h = Head loss due to vapor flow through perforations, in. liquid; or

= Dry tray pressure drop, in. liquid

h_L = Clear liquid head, m

h_l = Depth of clear liquid on tray, inches; (or m); or = Aerated tray, liquid pressure drop or equivalent clear liquid on tray, in. tray liquid

h_{li} = Height of clear liquid on inlet side of tray. in.

h_{lo} = Height of clear liquid at overflow weir, in.

h_n = Depth of notch in weir, in.; or = Head in the back of downcomer, in. (usually negligible)

h_{ow} = Height of liquid crest over flat weir; or measured from weir (straight or circular); or from bottom of notches (v-notch weir), in.

= Height of liquid above bottom of notch in notched weir, in.

h_{op} = H_{op} = Valve lift, i.e., distance between bottom of a valve and top of the tray deck, in.

h_{pc} = Cap assembly partial pressure drop, including drop through riser, reversal, annulus, slots, in. liquid

h_r = Pressure drop through risers, in. liquid

h_{ra} = Pressure drop through reversal and annulus, in. liquid

h_s = Slot opening, or pressure drop through slot, in. liquid

h'_s = Pressure drop through dry slots, in. liquid
h_{sl} = Static liquid seal on sieve tray, in. liquid
h_{ss} = Static slot seal, in.
h_{sr} = Height of cap shroud ring, in.
h_t = Total vapor pressure drop per tray, in. liquid (wet tray)
h_{udc} = Head loss due to the underflow clearance, in.
h_v = Maximum vertical travel of a valve on a valve tray, metric
h_w = Height of weir above tray; floor (to top of flat weir, or bottom of notch in notched weir), in.
h_{WT} = Wet tray head loss, in. liquid
K_c = Constant for Bolles' partial bubble cap pressure drop equation, Figure 13-52, or
= Loss coefficient, valve closed, $(sec)^2(in.)/(ft^2)$
L = L′ = Liquid flow, lb/hr or lb/sec (or m^3/hr/m weir length); or
= Liquid rate, lb/hr.(ft^2 active plate/tray)
L_{bc} = Liquid mass velocity, lb/hr-ft^2- based on superficial cross section of column
L/V = Internal reflux ratio, dimensionless
L_{Wi} = Weir length, in.
L_g = Liquid flow rate, gpm = Q
l_f = Liquid flow rate, ft^3/sec
l_{fw} = L_w = Total flow width across tray normal to flow, ft
l_w = Length of straight weir, ft
= l_{fp} = Length of liquid flow path, ft
m_{vap} = Vapor rate, lb/sec
m = Exponent in CCFL correlation, or Equation 13-114, equals approx. 1.0
N = Total number of actual trays in tower
N_c = Number of caps per tray

N_s = Number of slots per bubble cap

N_v = Valve density, number of valves per ft^2; or

= Number of valve units on a valve tray

n = Depth of notches in weir, in. or

= Exponent defined by Equations (13-72) and (13-114)

ΔP = Dry tray pressure drop for 50% cut baffles, in. liquid per baffle; or

= Actual tray pressure drop, in. liquid

P_r = Prandtl number, dimensionless

P_v = Fractional opening in the circumference or a valve; or, P_l

Q = Liquid load, gpm = L_g

R_h = Hydraulic radius for froth cross flow, ft

R_s = Ratio of top to bottom widths of trapezoidal slot, bubble cap dimensions

R_v = Vapor distribution ratio, dimensionless

R_{vw} = Ratio valve weight with legs/valve weight without legs, dimensionless, see Table 13-12

Re_h = Reynolds Number modulus

R_{eh} = Reynolds modulus for friction cross-flow

Δ'_r = Liquid gradient per row of caps, uncorrected, in.

S′ = Effective tray spacing, distance between top of foam, froth, or bubbles and tray above, in. Note: for Hunt's relation, S′ = tray spacing minus 2.5 h_c

S_c = Schmidt number, dimensionless

S′ = Same as S′, except unit, ft.

S_t = S = Tray spacing (actual), in. ft, m

S_{ts} = Tray spacing, ft.

s = Cap skirt clearance between cap and tray floor, in.

T_v = Metal thickness of valve, in.

t_w = Liquid throw over weir, in.

U = Superficial vapor velocity, m/sec

U_N = Vapor linear velocity based on net area for de-entrainment usually tower crosssection minus one downcomer, ft/sec
$U_a = V_a$ = Vapor velocity based on active area, A_a, ft/sec
V = Total vapor flow through tray or tower, ft^3 /sec
V′ = Internal vapor flow, lb/hr or lbs/sec, Equation 13-82
V_G = Superficial gas velocity in channel (not tower), ft/sec
V_{load} = Vapor load corrected for density, ft/sec
V_m = Maximum allowable vapor load per tray, ft^3/sec
v_c = Superficial vapor velocity in tower, ft/sec (based on tower crosssection)
v_d = Design hole vapor velocity, ft/sec; or
= Downcomer velocity, ft/sec
v_{du} = Velocity of liquid flowing between segmental downcomer and inlet weir, ft/sec
v_f = Vapor velocity through equivalent net tray area, based on tower area minus twice down comer area, ft/sec; also = Velocity of froth cross flow, ft/sec
v'_f = Velocity of froth, ft/sec
v_{flood} = Gas superficial velocity based on tray net area, A_n, ft/sec
v_h = Vapor velocity through valve hole, ft/sec
$v_{pt} = v_o$ = Vapor velocity through holes, ft/sec
$v_w = U_w$ = Linear velocity of gas based on window area, ft/ sec
v_{om} = Minimum velocity through holes at weep point, ft/sec
W = Maximum allowable mass velocity through column using bubble cap trays, lb/(hr) (ft^2 tower cross section)
W_e = Liquid entrainment mass velocity, lb entrainment/(min) (ft^2), based on net tray area of tower minus twice downcomer area
W'_e = Assumed allowable liquid entrainment mass velocity derived from assumed allowable loss mols liquid/mol vapor, lb/hr (ft^2), based on net tray areas same as for W_e

$W*_e$ = Liquid entrainment mass velocity corrected for liquid properties and plate spacing, lb entrainment/(hr) (ft^2) , based on net tray area as for W_e

w_l = weir length, in.

w_s = width of slot (rectangular), in.

z = Characteristic length in CCFL model, ft

Greek Symbols

α = Relative volatility, dimensionless

α_d = Mean aeration factor of froth (dimensionless)

= Relative froth density, h_L/h_f

β = Fraction perforated or open hole area in perforated area of tray (not fraction hole area in tower area); or

= Aeration factor, f, dimensionless

λ = Slope of equilibrium line/slope of operating line

Δ = Liquid gradient (corrected) for tray or tray section, in.

= Uncorrected liquid gradient for tray or tray section, in,

Δ′r = Liquid gradient per row of caps, uncorrected, in.

ϕ = Relative froth density, ratio of froth density to clear liquid density

ε = Eddy kinematic viscosity, m^2/sec (assumed equal to eddy diffusivity; see Ref. 2)

θ = Time to drain tower, min

μ = Viscosity of liquid at tower temperature, centipoise, cP

μ_l = Viscosity of liquid, lb/fr-sec

π = Pi = 3.14

Ψ = Entrainment expressed as fraction of gross downflow

ρ = Liquid density at temperature of tower, gm/cc

ρ_L = Liquid density, lb/ft^3 or kg/m^3

ρ_v = Vapor density, lb/ft^3 or kg/m^3
ρ_{vm} = Valve metal density, lb/ft^3
σ = Surface tension of liquid, dynes/cm

Subscripts

F = Flood = At flood point
g = G = Gas
H_2O = H = h = Water
L = l = Liquid
OG = og = Overall (gas concentration basis)
Vap = vap = Vapor

Chapter 14. Packed Towers

Packed towers are used increasingly in a variety of applications in the chemical process industries. The reasons for this are an increase in the availability of design information, the evolution of higher-capacity and higher-efficiency mass transfer packings, and improvements in distributors and support plates. Today, packings can be considered for most services where high numbers of theoretical stages are required for mass transfer.

Most of the fractionating columns in gas-processing plants that were previously equipped with trays can now be filled with packings. Packings, such as slotted rings in metal and plastic saddles, offer large improvements in capacity for absorption and heat transfer applications. An example of this is the use of slotted rings in a gas-plant demethanizer, which has resulted in a significant capacity improvement over trays.

Packed towers are used increasingly in small fractionating towers and absorbers where access is restricted. Their use is also advantageous where corrosion control is essential. Because of liquid-distribution concerns, packing still is usually not recommended for large-diameter columns, or where a high turndown ratio is required. All packings tend to spread liquid as it flows downwards. This characteristic is adequate for dispersing the liquid between distribution points, but not to correct poor distribution across a tower without considerable loss of efficiency. Studies of liquid spread in tower packing have been reported by Hoek [158]. Figure 14-1A shows a crosssection of a typical packed tower, Figure 14-1B illustrates a typical packed tower with internals for improved distillation. Figure 14-1C and Figure 14-1D show a photograph of a packed column, and a schematic of the

pilot plant packed column respectively. The shell is usually cylindrical, although square wooden, light metal, or reinforced plastic towers are used. The basic unit consists of:

1. shell

2. packing (one or more sections)
3. packing support(s)
4. liquid distributor(s)

5. intermediate supports and redistributors

6. gas and liquid entrance and exit nozzles.

Figure 14-1A

Crosssection of typical packed tower.

Figure 14-1B

Typical packed tower with internals for improved distillation.

Used by permission of Jaeger Products Inc., Bull. 1100.

Figure 14-1C

Photograph of a pilot plant absorption column

(courtesy of Armfield, U.K.).

Schematics of a pilot plant absorption column

(courtesy of Armfield, U.K.).

Many of the mechanical aspects of tower construction and assembly have an influence on the design and interpretation of tower performance. Every effort should be made to increase the effectiveness of contact between the process streams, and to reduce losses by entrainment or wall effects whilst minimizing pressure drop. At the same time the design must be consistent with the economics dictated by the process and type of construction.

14.1. Shell

The shell may be of metal (steel, alloy or non-ferrous), plastic, wood or some combination, which may require the addition of liners or inner layers of rubber, plastic or brick. The mechanical problems of attaching inner nozzles, supports and brick require considerable attention indesign that is not an integral part of sizing the equipment. Figure 14-2AFigure 14-2B and Figure 14-2C show a typical large, steel brick-membrane lined tower with corbeled brick support locations. In these towers, temperature and/or corrosive conditions usually dictate the composition of internal lining, and the requirements for the proper acid-(or alkali-) proof cements.

Figure 14-2A

Crosssection of membrane and brick-lined parked tower. Depending on tower diameter, certain dimensions should be modified, particularly parking. Also see

Figure 14-2B and Figure 14-2C.

Figure 14-2B

Crosssection of membrane and brick-lined packed tower. Depending on tower diameter, certain dimensions should be modified, particularly packing.

Figure 14-2C

Crosssection of membrane and brick-lined packed tower. Depending on tower diameter, certain dimensions should be modified, particularly packing.

Ceramic, plastic and other nonmetal tower shells are used quite often (Figure 14-3Figure 14-4 and Figure 14-5). It is important in ceramic construction to consider that the main inlet or outlet nozzles, or any other large connections, should be oriented 90° to each other in order to reduce the possibility of cracking the walls, as most cracks go one-half diameter. Preferably there should only be one nozzle in any one horizontal plane. The nozzles should never carry any piping or other stress load.

Figure 14-3

Bell and spigot ceramic tower: Used by permission General Ceramics and Steatite Corp.

Figure 14-4

Physical dimensions of stoneware tower sections, bell and spigot.

Used permission of General Ceramics and Steatite Corp.

Figure 14-5

Typical reinforce plastic tower construction; with dimension used by permission of Haveg Corp., Bull. F-7.

The bell and spigot type tower, shown in Figure 14-3 and Figure 14-4, is satisfactory for 2 to 2.5 psi in 12-in. dia. to 30-in. dia. towers when the joints are packed with chevron-type caulking compound. For an operating pressure of 5 psi in 18-in. through 48-in. dia., use of non-

asbestos and silicate cement. Special hold-down packing gland-type rings will allow this pressure to be raised slightly. Porcelain towers should be used for higher pressures rather than the weaker stoneware.

The rate of heating or cooling a stoneware or porcelain tower should not exceed 15°F/min.

14.2. Random Packing

The distributor and packing define the performance of this equipment. Their proper selection entails an understanding of packing operational characteristics, and the effect on performance of the points of significant physical difference between the various types.

Good progress has been made in the past decade in the development of packing types for difficult and wide ranging process applications (see: Websites in bibliography). These types include:

1. *Random particle* packings are discrete, individually shaped particles designed to provide contacting surfaces between (normally) downflowing liquid and upflowing vapor/gas. The degree of effectiveness of the various shapes varies along with the mass pressure drop through the packed bed. Usually these particles are "dumped" into the column (tower) and allowed to float gently to their free-fall resting position in a column full of water. Some shapes and sizes are not installed using water, but dumped in using a special "sock" that allows the particles to be lowered without a damaging free fall. Sometimes large particles are hand set "dry" into position to fill the tower tightly, See Figure 14-1A. The vapor-liquid performance is different for each method of loading, and appropriate data must be available to properly size the tower. The random "dumped" (wet or dry) method is the technique usually used in most published data (see Figure 14-6AFigure 14-6BFigure 14-6CFigure 14-6DFigure 14-6EFigure 14-6FFigure 14-6GFigure 14-6HFigure 14-6IFigure

14-6JFigure 14-6KFigure 14-6LFigure 14-6MFigure 14-6N(a)Figure 14-6N(b)Figure 14-6OFigure 14-6PFigure 14-6QFigure 14-6RFigure 14-6SFigure 14-6TFigure 14-6UFigure 14-6VFigure 14-6W and Figure 14-6X).

Figure 14-6A

Various packing shapes (ceramic).

Figure 14-6B

Raschig rings (ceramic, carbon, metal).

Figure 14-6C

Intalox® saddles (ceramic), dumped.

Used by permission of Norton Chemical Process Products Corp.

Figure 14-6D

Berl saddles (ceramic), dumped.

Figure 14-6E

Pall rings (metal), dumped.

Figure 14-6F

Koch Flexring (plastic).

Used by permission of Koch Engineering Co., Inc., Bull. PFR-1.

Figure 14-6G

Norton Hy-Pak® (metal).

Used by permission of Norton Chemical Process Products Co., Inc. Bull. DC-11 and MTP-1 (4/94).

Figure 14-6H

Plastic pall ring. Note: Glitsch Ballast Ring® and others are quite similar.

Used by permission of Norton Chemical Process Products Co., Inc. Bull. DC-11 and PTP-1 (11/87).

Figure 14-6I

Metal pall ring. Note: Glitsch Ballast Ring® and Koch Engineering Flexring® are quite similar.

Used by permission of Norton Chemical Process Products Co., Inc. Bull. N-60D and Bull. MTP-1 (4/94).

Figure 14-6J

Metal Intalox (IMTP®).

Used by permission of Norton Chemical Process Products Co., Inc. Bull. 1HP-1 (12/91).

Figure 14-6K

Nutter Ring™ (metal random packing).

Used by permission of Nutter Engineering, Harsco Corp., Bull. NR-2.

Figure 14-6L

Metal Top-Pak® low liquid/high gas flow and high heat transfer packing.

Used by permission of Vereinigte Füllkörper – Fabriken GmBH & Co. Ransbach Baumbach, Germany.

Figure 14-6M

Metal VSP® high capacity packing.

Used by permission of Vereinigte Füllkörper – Fabriken GmBH & Co. Ransbach Baumbach, Germany.

Figure 14-6N(a) Koch Metal HcKp™ random packing.

Used by permission of Koch Engineering Co., Inc. Bull. KRP-2.

Figure 14-6N(b) Fleximax™ high-performance random metal packing; available in two sizes and reported by manufacturer to be a fourth generation random packing.

Used by permission of Koch Engineering Co., Inc. Bull. KFM-1.

Figure 14-6O

Koch Metal Flexring® ; available in ⅝ –in.–3 ½ in. diameter.

Used by permission of Koch Engineering Co., Inc. Bull. KRP-2.

Figure 14-6P

Metal Chempak® packing.

Used by permission of Chem-Pro Equipment Corp., licensed from Dr. Max Leva.

Figure 14-6Q

Plastic Nor-Pac® packing; fabricated of most corrosion-resistant plastics.

Used by permission: NSW Corporation.

Figure 14-6R

Super Intalox® Saddles. Note: Glitsch Ballast Saddle® and koch Engineering Flexi saddle® are quite similar.

Used by permission of Norton Chemical Process Products Corp., Bull. N-60D.

Figure 14-6S

Koch Flexisaddles™ (plastic); available in 1-in.–3-in.

Used by permission of Koch Engineering Co., Inc., Bull. KRP-2.

Figure 14-6T

(Top) Cascade® Mini-Ring, (metal and plastic). Originally used by permission of Mass Transfer, Inc., now Glitsch, Inc., (middle and bottom) Elevation and plan views of Ballast® rings (right) and Cascade Mini-Rings (left). Note how high aspect ratio of former permits occlusion of interior surfaces. Low aspect ratio of Cascade Mini-Rings, on the other hand, favors orientation that exposes internal surfaces for excellent film formation, intimate mixing, and gas–liquid contact.

Used by permission of Glitsch, Inc. Bull. 345.

Figure 14-6U

Norton Intalox® high-performance Snowflake® packing (plastic).

Used by permission of Norton Chemical Process Products Corp., Bull. 1SPP-1R.

Figure 14-6V

Jaeger Tri-Packs® high-performance fabricated of corrosion-resistant plastics.

Used by permission of Jaeger Products, Inc.

Figure 14-6W

Rauschert Hiflow® high-performance rings and saddles packing (plastic and metal).

Used by permission of Rauschert Industries, Inc., div. of Rauschert GmbH & Co., KG; Paul-Rauschert-Str., 6; D-96349 Steinwiesen, Germany.

Figure 14-6X

Teller rosette packing (plastic).

Used by permission of Hawshaw Chemical Co., and Dr. A. J. Teller.

2. *Structured packing*, which is offered by several manufacturers, is usually composed of pack "pads" fabricated by shaping/crimping, bending, rolling, *etc*. sheets of thin gauge metal or wire, (see Figure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EEFigure 14-6FFFigure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MMFigure 14-6NN and Figure 14-6OO). Some "pads" or packs are formed using various plastics material, selected to be resistant to the fluid services involved.

Figure 14-6Y

Panapak formed/structured packing for a 6-ft-diameter column (original design).

Used by permission of Packed Column Corp.

Figure 14-6Z

Panapak packing for a 6-ft-diameter column, latest design.

Used by permission of Packed Column Corp.

Figure 14-6AA Cartridges 8½-ft high of 32 in. I.D. Spraypak Packing.

Used by permission of Denholme, Inc. Bull. No. 8.

Figure 14-6BB

Goodloe packing.

Used by permission originally of Packed Column Corp., now under license to Glitsch, Inc.

Figure 14-6CC

York-Twist™ structured tower packing. Note insert showing weave and sections fabricated to fit through manway for larger towers.

Used by permission of Otto H. York Co., Inc., a division of Glitsch, Inc.

Figure 14-6DD

ACS knitted mesh structured packing.

Used by permission of ACS Industries, Inc., Separations Technology Group, Bull. B-129 (1992).

Figure 14-6EE

Koch/Sulzer® packing.

Used by permission of Koch Engineering Co., Inc., Bull. KS-2.

Figure 14-6FF

Liquid and gas mixing in Koch/Sulzer® packing elements.

Used by permission of Koch Engineering Co., Inc., Bull. KS-2.

Figure 14-6GG

Koch Flexipac® structured packing.

Used by permission of Koch Engineering Co., Inc., Bull. KFP-4.

Figure 14-6HH

Metal Max-Pak™ structured packing fabricated from corrugated sheets.

Used by permission of Jaeger Products, Inc. Bull. 500.

Figure 14-6II Intalox® high-performance structured packing.

Used by permission of Norton Chemical Process Products Corp., Bull. IS-2T.

Figure 14-6JJ

Intalox® high-performance wire gauze structured packing.

Used by permission of Norton Chemical Process Products Corp., Bull. WG-1.

Figure 14-6KK

Glitsch Gempak® structured packing series designs. Series AT – Smooth surface with small crossover points on narrow rectangular pitch. A general purpose style, especially suitable for low wetting rates and high vacuum applications. Series AS – A general purpose style suited for high wetting and heat transfer applications. Both AT and AS series are particularly suited for refinery applications and can be used in the wash zones of atmospheric and vacuum crude towers, or FCC main fractionators. Series AL – Metal surface with louvers in a repetitive pattern. Especially suited for low wetting rates and chemical applications. Series BG – A wire gauze packing, crimped to 60° from the horizontal. Recommended for high efficiency applications at very low wetting rates (clean service).

Used by permission of Glitsch, Inc. Bull. 5140 (1993).

Figure 14-6LL

Montz-Nutter high-efficiency structured packing assembly (Type B-1™ with wiper bands and seal strip).

Used by permission of Nutter Engineering, Harsco Corp., Bull. B-1.

Figure 14-6MM

Nutter BHS™ structured packing with expanded metal texture, which maintains high mass transfer efficiency across all operating conditions. Nutter Engineering designs and manufactures BSH in North America under exclusive license from Julius Montz GmbH of Germany.

Used by permission of Nutter Engineering, Harsco Corp.

Figure 14-6NN

Koch Ceramic Flexeramic® structured packing of stoneware or ceramic fabrication.

Used by permission of Koch Engineering Co., Inc., Bull. KCP-2.

Figure 14-6OO

Installation of Ceramic Flexeramic® structured packing of stoneware or ceramic.

Used by permission of Koch Engineering Co., Inc., Bull. KCP-2.

3. *Grid* packing is probably the newest packing type. It is lower in pressure drop, and has higher capacity, but lower efficiency than the other types. (See Figure 14-6PPFigure 14-6QQFigure 14-6RRFigure 14-6SSFigure 14-6TT and Figure 14-6UU).

Figure 14-6PP

Nutter Snap-Grid™ grid packing as assembled.

Used by permission of Nutter Engineering, Harsco Corp., Bull. CSG-1.

Figure 14-6QQ

Nutter Snap-Grid™ high-capacity grid packing with good structural integrity (interlocking) and reduced fouling.

Used by permission of Nutter Engineering, Harsco Corp., Bull. CSG-2.

Figure 14-6RR

Intalox® Grid-type packing; useful for fouling, solids and heat transfer; high-strength, low-pressure drop, high capacity, better efficiency than most structured and random packing.

Used by permission of Norton Chemical Process Products Corp., Bull. IG-1.

Figure 14-6SS

Flexigrid® Style 3 high-efficiency structured packing.

Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

Figure 14-6TT

Flexigrid® Style 2 high-efficiency structured packing.

Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

Figure 14-6UU

Glitsch-grid™ packing, Style EF-25A.

Used by permission of Glitsch, Inc., Bull. 163L84B.

The types and corresponding physical data for packing are given in Figure 14-6AFigure 14-6BFigure 14-6CFigure 14-6DFigure 14-6EFigure 14-6FFigure 14-6GFigure 14-6HFigure 14-6IFigure 14-6JFigure 14-6KFigure 14-6LFigure 14-6MFigure 14-6N(a)Figure 14-6N(b)Figure 14-6OFigure 14-6PFigure 14-6QFigure 14-6RFigure 14-6SFigure 14-6TFigure 14-6UFigure 14-6VFigure 14-6WFigure 14-6XFigure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EEFigure 14-6FFFigure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MMFigure 14-6NNFigure 14-6OOFigure 14-6PPFigure 14-6QQFigure 14-6RRFigure 14-6SSFigure 14-6TT and Figure 14-6UU and Table 14-1Table 14-2Table 14-3Table 14-4Table 14-5Table 14-6ATable 14-6BTable 14-7Table 14-8Table 14-9Table 14-10Table 14-11Table 14-12Table 14-13Table 14-14 and Table 14-15. Evaluation of these materials in various conditions of service is provided later. However, Table 14-16 outlines packing service applications and Table 14-17 summarizes the usual applications of each packing type.

Table 14-1 Chemical Stoneware and Chemical Porcelain Raschig Rings.

Normal size, in.	Wall thickness, in.
½	
⅝	
1	⅛
1¼	
1½	¼
2	¼
3	⅜
+3(D)	⅜
+3(S)	⅜
4	½
⅝, 2½, 4, and 6 in. available on special order. The 3-in., 4-in., and 6-in. O.D. size surfaces. The 3-in., 4-in., and 6-in. sizes can be made in lengths up to 12 in., on s	
**Porcelain Rings are about 5% heavier than Stoneware. These weights are the a +Data for stacked arrangement. “D” indicates diamond pattern. “S” indicates squ Original by permission, M. Leva, Ref. 40 and Bulletin TP-54, U.S. Stoneware Co of Norton Chemical Process Products Corp.	

From: M. Leva, U.S. Stoneware Co., Akron, Ohio, Ref. 40, and Bulletin CP-25 York, N.Y.; updated 12/93 by permission Norton Chemical Process Prod

Table 14-2 Carbon Raschig Rings.

Nominal		O.D.	Approx. avg.	Approx avg.

Nominal size, in.	Wall thickness, in.	and length, in.	number per Ft^3	weight per ft^3 lb
¼		¼	85,000	46
½		½	10,600	27
¾	⅛	¾	3,150	34
1	⅛	1	1,330	27
1¼		1¼	678	31
1½	¼	1½	390	34
2	¼	2	166–157*	27
3		3	50–44*	23
3(D)		3	74	49.5
3(S)		3	64	43
*Varies with supplier.				

Table 14-3 Metal Raschig Rings.

Original by permission, M. Leva,[40] and U.S. Stoneware Co., Akron, OH.; updated 9/88 by permission of Norton Chemical Process Products Corp., Bull. MTP-1 and TP-78.

Nominal size, in.	Wall thickness, in.
¼	
⅜**	
½**	
½**	
⅝**	

⅞	
¾**	
¾**	
1**	
1**	
1¼**	
1½**	
2**	
3**	
3(D)	
3(S)	
Usually metal Raschig rings are made with fitted butted-joints.	
Note: Sizes availability varies with manufacturers.	
*Based upon Carbon Steel Rings; other weights are: Stainless Steel 105%; Copp	
**Usual commercial sizes.	

Table 14-4 Metal Lessing Rings.

Size, in.	Wall thickness, in.	O.D. and length, in.	Approximate number per ft^3	Appro weig ft^3,
¼		¼	81,840	195
⅜		⅜	25,110	114
½		½	10,974	100
½		½	10,230	172

½		½	10,250	172
⅝		⅝	6,789	86
⅝		⅝	6,510	150
¾		¾	3,171	71
¾		¾	2,967	130
1		1	1,339	52
1		1	1,251	95
1¼		1¼	674	81
1½		1½	391	65
1¾		1¾	246	58
2		2	167	49
All figures are on a dumped basis. Metal Lessing Rings are also made in stainles				
* Weights shown are for carbon steel. By permission: U.S. Stoneware Co. Bul. T Products Corp., Stow, OH. [5] except column 8, [40].				

Table 14-5 Ceramic Lessing Rings.

From M. Leva, U.S. Stoneware Co., [40], by permission, now, Norton Chemica Corp., Stow, OH.

O.D. and length, in.	Wall thickness, in.	Approx. no. of rings per ft^3, dumped	Approx. wt. per ft^3 dumped	Percent free gas space
1	⅛	1,300	50	66
1¼		650	56	62
1½	¼	350	58	60
2	⅜	150	49	68

Table 14-6A Chemical Stoneware and Ceramic Berl Saddle Packings.

From M. Leva, U.S. Stoneware Co., [40], by permission.

	Approx.	Approx.	Approx.	Percent	

Nominal size	Approx. average number/ft^3	Approx. wt./ft^3, lb	Approx. average surface area ft^2/ft^3	Percent free gas space*	Equivalent spher. diam. dp, in.
¼	113,000	56	274	60–67	0.23
½	16,200	54	142	63	0.42
¾	5,000	48	82	66	0.58
1	2,200	45	76	69–70	0.76
1½	580	38	44	73–75	1.10
2	250	40	32	72–75	1.55
*Updated by permission Jaeger Products, Inc.					

Table 14-6B Steel* Berl Saddles. Courtesy Maurice A. Knight Co., Bulletin No. 11, now, by permission Koch Engineering Co., Inc.

1	2,500	87	85	83
1½	825	60	58	88
*Other metals available.				

Table 14-7 Ceramic Intalox®* Saddles. Orig. By permission U.S. Stoneware Co., Akron, Ohio, Ref. 5, except column 6, from Ref. 40; updated 9/88 by permission of Norton Chemical Process Products Corp., Inc., Bull. CTP-1, 9/88.

Nominal size, in.	Approximate number per ft^3	Approximate weight per ft^3, lb	Approximate surface area ft^2/ft^3	Percent free gas space	Equivalent spherical diameter dp, in.
¼	117,500	54	300	64	0.20
⅜	50,000	50	—	67	—

½	17,000	46	190	69	0.32
¾	6,200	44	102	71	0.48
1	2,100	42	78	72	0.68
1½	660	39	59.5	74	0.96
2	240	37	36	75	1.38
3	52	37	—	75	—
Data shown applies to Intalox Saddle made either from chemical stoneware or chemical porcelain. Weights per cubic foot are based on chemical porcelain. Chemical stoneware Intalox Saddles will weigh approximately 5% less.					
*Trade name Norton Co.					

Table 14-8 Ceramic Cross-Par

			Number of ringers per ft^3	
Outside diameter in.	**Outside length in.**	**Wall thickness, in.**	**D setting**	**S setting**
3*	3	⅜	74	50*–64
4	3	½	41	36
4	4		31	27
6	4	⅝	14	12
6	6	⅝	9	8
Also made with outer surfaces ribbed or corrugated, and in lengths up to 12 in. R on special order. Porcelain rings weigh about 5% more than above. For D and S Stoneware Bull. TP-54, and Maurice A. Knight Co. (now, Koch Engineering Co. 2/89 by permission of Norton Chemical Process Products Corp. Bull. WG-1.				
For 3-in. size only;				
*indicates dumped rings only; all others stacked diamond (D) or square pattern s				

*Indicates dumped rings only; all others stacked diamond (D) or square pattern s

Table 14-9 Ceramic Spiral

				Numbe rings p	
	Outside diam. in	**Outside length in.**	**Wall thickness in.**	**D setting**	**s**
	3¼−3	3		63–74	5
Single	4	4	⅝	31	2
Spiral	6	6	½	9	8
	3¼	3		63	5
Double	4	4	⅝	31	2
Spiral	6	6	½	9	8
	3¼	3		63	5
Triple	4	4	⅝	31	2
Spiral	6	6	½	9	8

*Basic data in table for U.S. Stoneware "Cyclohelix" spiral packing, Bul. TP 54, Maurice A. Knight Co. (now, Koch Engineering Co., Inc.) Bulletin No. 11, by pe

Table 14-10 Pall Rings—Metal

By permission, Norton Chemical Process Products Corp., TP-78 and PR-16 and MTP-1 (9/88); other manufacturer's data are equivalent.

Size O.D. & length, in.	Wall[†] thickness, in.	Approx. number ft^3	Approx.* weight, lb/ft^3	Surface area ft^2/ft^3	Percent free gas space
⅝	0.018	5,800	37	104	93
1	0.024	1,400	30	63	94

1	0.024	1,400	30	65	94
1½	0.030	375	24	39	95
2	0.036	165	22	31	96
3½	0.048	33	17	20	97
		Plastic[‡]			
⅝	—	6,060	5.95		87
1	—	1,430	4.4		90
1½	—	420	4.35		91
2	—	175	3.85		92
3½	—	33	3.45		92
[†]Standard gauge carbon steel, approximate					
[‡]Weights referenced to polypropylene; other plastics available, high density polyethylene, glass reinforced polypropylene and fluorinated vinyls.					

Table 14-11 Teller Rosette (Tellerette) Plastic[*].

Nominal size, in.	No. units per ft^3	Weight per ft^3, lb	Surface area ft^2/ft^3	Percent free gas space
1	1125	10	76	83
*Harshaw Chemical Co. “Tellerette” bulletin, and Dr. A. J. Teller.				

Table 14-12 Super Intalox® Saddles Ceramic[†].

By permission, Norton Chemical Process Products Corp., Bull. SI-72 and Bull. PTP-1; other manufacturer's data are equivalent.

Size no. designation	Approximate number/ft^3	Weight[‡] lb/ft^3	Surface area ft^2/ft^3	Percent free gas space
1	1,500	35	76	77
2	180	38	32	75
		Plastic		
1	1,500	5.85	63	90
2	180	3.75	32	92

2	180	3.75	33	93
3	42	3.00	27	94
†Also available in polypropylene (including glass reinforced); high density polyethylene, rigid PVC, fluorinated vinyls.				
‡Weights for polypropylene; others are times PP: 1.03 for high density polyethylene: 1.54 for PVC: 1.87 to 1.95 for fluorinated vinyls.				

Table 14-13 Hy-Pak™* Metal Packing.

Size no. designation	Approximate number/ft^3	Weight, ** lb/ft^3	Percent free gas space
1	850	19	97
1½	270	—	97
2	105	14	98
3	30	13	98
*By permission, Norton Chemical Products Corp., Bull. HY-30®, dated 9/88 reg. trademark.			
**Weight for standard gauge carbon steel, available in most other common metals.			

Table 14-14 Chempak® Metal Packing*.

1	950	19	96
*By permission Chem-Pro Equipment Corp., Licensed from Dr. Max Levo, Bull. 702. Weight is for carbon steel.			

Table 14-15 Grid Tile (Drip Point)*.

Table 14-16 Packing Service Application.

Packing material	General service application	Remarks
Glazed and unglazed, Porcelain or Chemical Stoneware	Neutral and acid conditions except hydrofluoric, solvents. Not good in hot caustic (above 70°F)	Unglazed usual type specified except special requirement of low adsorption on surface. Special ceramics available for mild caustic. Porcelain stronger and more resistant than stoneware
Carbon	Hot alkali, all acids except nitric, no oxidizing atmospheres	Stand thermal shock, low cubic weight
Plastic	Alkali, salts, aqueous and acids depending on resin	Light weight
Steel and other light gauge metals	Hot alkali for steel, other service to suit metals	May be heavier than ceramic, more expensive

Table 14-17 Packing Type Application.

Packing	Application features
Raschig Rings	Earliest type, usually cheaper per unit cost, but sometimes less efficient than others. Available in widest variety of materials to fit service. Very sound structurally. Usually packed by dumping wet or dry, with larger 4–6 in. sizes sometimes hand stacked. Wall thickness varies between manufacturers, also some dimensions; available surface changes with wall thickness. Produce considerable side thrust on tower. Usually has more internal liquid

	channeling, and directs more liquid to walls of tower. Low efficiency
Berl Saddles	More efficient than Raschig Rings in most applications, but more costly. Packing nests together and creates “tight” spots in bed which promotes channeling but not as much as Raschig rings. Do not produce as much side thrust; has lower HTU and unit pressure drops with higher flooding point than Raschig rings. Easier to break in bed than Raschig rings
Intalox Saddles[1] And Other Saddle-Designs	One of most efficient packings, but more costly. Very little tendency or ability to nest and block areas of bed. Gives fairly uniform bed. Higher flooding limits and lower pressure drop than Raschig rings or Berl saddles; lower HTU values for most common systems. Easier to break in bed than Raschig rings, as ceramic
Pall Rings[2]	Lower pressure drop (less than half) than Raschig rings, also lower HTU (in some systems also lower than Berl saddles), higher flooding limit. Good liquid distribution, high capacity. Considerable side thrust on column wall. Available in metal, plastic and ceramic
Metal Intalox®[1] Hy-Pak[1] Chempak[7]	High efficiency, low pressure drop, reportedly good for distillations
Spiral Rings	Usually installed as stacked, taking advantage of internal whirl of gas-liquid and offering extra contact surface over Raschig ring, Lessing rings or cross partition rings. Available in single, double and triple internal spiral designs. Higher pressure drop. Wide variety of performance data not available
Lessing Rings	Not much performance data available, but in general slightly better than Raschig ring, pressure drop slightly higher. High side wall thrust
Cross-Partition Rings	Usually used stacked, and as first layers on support grids for smaller packing above. Pressure drop relatively low, channeling reduced for comparative stacked packings. No side wall thrust
Grid Tile	Available with plain side and bottom or serrated sides and drip-point bottom. Used stacked only. Also used as support layer for dumped packings. Self supporting, no side thrust. Pressure drop lower than most dumped packings and some stacked, lower than

	some ¼-in. × 1-in. and ¼-in × 2-in. wood grids, but greater than larger wood grids. Some HTU values compare with those using 1-inch Raschig rings
Teller Rosette (Tellerette)	Available in plastic, lower pressure drop and HTU values, higher flooding limits than Raschig rings or Berl saddles. Very low unit weight, low side thrust
Spraypak[3]	Compared more with tray type performance than other packing materials. Usually used in large diameter towers, above about 24-in. dia., but smaller to 10-in. dia. available. Metal only
Panapak[4]	Available in metal only, compared more with tray type performance than other packing materials. About same HETP as Spraypak for available data. Used in towers 24 in. and larger. Shows some performance advantage over bubble cap trays up to 75 psia in fractionation service, but reduced advantages above this pressure or in vacuum service
Stedman Packing	Available in metal only, usually used in batch and continuous distillation in small diameter columns not exceeding 24-in. dia. High fractionation ability per unit height, best suited for laboratory work. Conical and triangular types available. Not much industrial data available
Sulzer, Flexipac, and similar	High efficiency, generally low pressure drop, well suited for distillation of clean systems, very low HETP.
Goodloe Packing[5] and Wire Mesh Packing	Available in metal and plastic, used in large and small towers for distillation, absorption, scrubbing, liquid extraction. High efficiency, low HETP, low pressure drop. Limited data available.
Cannon Packing	Available in metal only, low pressure drop, low HETP, flooding limit probably higher than Raschig rings. Not much literature data available. Used mostly in small laboratory or semi-plant studies
Wood Grids	Very low pressure drop, low efficiency of contact, high HETP or HTU, best used in atmospheric towers of square or rectangular shape. Very low cost
Poly Grid[6]	Plastic packing of very low pressure drop, developed for water-air cooling tower applications

Compiled from literature:
[1]Trade name of Norton Chemical Process Products Corp.
[2]Introduced by Badische Anilin and Sodafabrik, Ludwigshafen am Rhein
[3]Trade name of Denholme Inc., Licensed by British Government
[4]Trade name of Packed Column Corp.
[5]Trade name of Packed Column Corp.
[6]Trade name of The Fluor Products Co.
[7]Trade name of Chem-Pro Equip. Corp.

14.2.1. Packing Supports

The packing support may be anything from cross-grid bars spaced so as to prevent fall-through of packing, to more refined, speciality units designed to direct the flow of gas and liquid. (See Figure 14-7AFigure 14-7BFigure 14-7CFigure 14-7DFigure 14-7E and Figure 14-7F.) Good tower performance is definitely linked to proper packing support. The net free flow crosssectional area of the support should be 65% (or larger) of the tower area, and greater than the free area of the packing itself. In addition, the effect of any free area “blocking” by the positioning of the packing on the support must be considered. To allow for this, every effort should be made to obtain as large a support-free area as possible, and yet remain consistent with the structural strength of the material being used. If this area is too restricted, liquid buildup will occur at the plate, reducing efficiency and increasing the pressure drop of the tower, which could lead to a flooding condition. A lot depends on the material of construction that the system requires; for example, carbon or graphite bar grids, brick grid piers, some steel grating grids and most rubber or plastic covered metal grids have inherently low free crosssectional areas. These may be less than 65% free area. Pressure drops through support plates, such as that shown in Figure 14-7D, are reported [82] not to exceed 0.3 in. water for most applications. Also see Figure 14-7AFigure 14-7BFigure 14-7CFigure 14-7E and Figure 14-7F. Fouling service can

create pressure drop problems.

Figure 14-7A

Bar-grid support.

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

Figure 14-7B

Perforated support plates.

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Process Products Corp.).

Figure 14-7C

Bar-grid support plate, typical details.

Figure 14-7D

Typical efficient metal and porcelain support plate designs for gas injection into packing. Designs also available in platics-FRP, polypropylene, PVC, *etc.* (also see Figure 14-7F).

Used by permission of Norton Chemical Process Products Corp.

Figure 14-7E

Reinforeced plastic support plate.

Used by permission of Haveg Corp., Bull. F-7. Note: All dimensions are in inches unless otherwise noted. Perforated plate free area should be increases when possible.

Figure 14-7F

Support plate for high gas/vapor flow (also see Figure 14-7D).

Used by permission of Jaeger Products, Inc.

In some large towers, the support grid is built up from supporting brick arches coming from the bottom (see Figure 14-2AFigure 14-2B and Figure 14-2C). Quite often in large towers, drip point grid tile is used as the supporting first layer, either as a support “plate” itself, or

as the support for other packing stacked on it (Table 14-15). This initial stacking of the first, and perhaps second courses of packing prevents the blocking of free area usually associated with dumping packing on support plates. The resulting net free area "balance" around the support grid or plate and its first two courses of packing (whether dumped or stacked) should be calculated in order to evaluate its effect on tower performance. Figure 14-1AFigure 14-1B and Figure 14-3 show a typical arrangement of several support plates.

The weight to be carried by the support plate is the sum of the weight of the packing, plus the weight of the flooded liquid volume of the packing voids, plus any pressure surges that might be imposed on the system. The effect of side thrust of the packing in reducing the dead packing load on the support should be ignored as it is an indeterminate figure. Normally, each support is required to support only the weights of the packed section directly above it, and not those separately supported above or below it. If any intermediate supports or redistributors are not separately supported on the tower wall but rest on the packing itself, the bottom support would carry the entire tower load of packing as mentioned, plus the weights of intermediate support and redistribution plates. This is not good practice as it complicates the packing and repacking of the tower as well as putting possibly excessively heavy loads on the bottom supports.

As a general rule, packing heights per support plate should not exceed 12 ft for Raschig rings, or 15–20 ft for most other packing shapes. Other types fit within these limits. Mechanical, vibrational and thermal shock loads can become important and sometimes take tower operation beyond these limits.

14.2.2. Liquid Distribution

Liquid distribution probably plays the most important part in the efficient operation of a packed tower. A good packing from the

process viewpoint can be reduced in effectiveness by poor liquid distribution across the top of its upper surface, or across the packing sections below any feed inlet(s) or reflux inlets. Poor distribution reduces the effective wetted packing area and promotes liquid channeling.

Final selection of the mechanism for distributing the liquid across the packing depends upon the size of the tower, type of packing (exposed surface, configuration), tendency of packing to divert liquid to tower walls, and materials of construction for distribution. Figure 14-8AFigure 14-8BFigure 14-8CFigure 14-8DFigure 14-8EFigure 14-8FFigure 14-8GFigure 14-8HFigure 14-8IFigure 14-8JFigure 14-8K and Figure 14-8L illustrate a number of available distribution types. Spray nozzles are used, but care must be taken in evaluating the fraction of the total liquid that hits the walls and never enters the packing. Full cone nozzles, with spray angles that keep most of the liquid on the center portion of the packing for initial contact, will perform quite well.

Figure 14-8A

Multilevel distributor with splash plate.

Figure 14-8B

Inlet pipe for distributor, Figure 14-8A.

Figure 14-8C

Distributor for intermediate level distribution.

Figure 14-8D

Weir-flow distributor.

Used by permission of U.S. Stoneware Co., Bull. TA-40 (now, Norton Chemical Process Products Corp.).

Figure 14-8E

Trough liquid distributor.

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

Figure 14-8F

Typical design inlet liquid distributor using holes (orifices) on underside of distribution pipes. Note: Number of side pipes adjusted to provide uniform entry per square foot of tower crosssection area. Holes at wall should be spaced on same basis.

Figure 14-8G

Koch/Sulzer® trough distributor (MDP), for medium to large diameter columns for structured packing (patented), one of several designs available.

Used by permission of Koch Engineering Co., Inc., Bull. KS-6.

Figure 14-8H

Orifice pan liquid distributor with round or rectangular chimneys and a flat floor sealed to vessel support ring. For a large vessel, it can be sectioned to pass through a manway.

Used by permission of Jaeger Products, Inc., Bull.-1100.

Figure 14-8I

Multipan MTS-109 two-stage liquid distributor.

Used by permission of Nutter Engineering, Harsco Corp., Bull. CN-4.

Figure 14-8J

MTS-109 Multipan two-stage liquid distributor for optimum liquid distribution for uniform flow; for random and structured packings for low to moderate liquid rates, less than 5 gpm/ft^2. Also used for redistributor.

Used by permission of Nutter Engineering, Harsco Corp., Bull. TI-1, under license from The Dow Chemical Co., protected by U.S. Patents No. 4,472,325; 4,808,350; 5,013,491.

Figure 14-8K

Spray nozzle distributor using full-cone wide-angle spray nozzles (depends on tower diameter). Not recommended for distillation applications, where the point-type distributors provide higher packing efficiency.

Used by permission of Nutter Engineering, Harsco Corp., Bull. TI-1

Figure 14-8L

Full cone spray nozzle as used for liquid distribution.

Used by permission of Spraying Systems, Co., Cat. 55.

Spray nozzle manufacturers have spray angle data for various pressures at the nozzle inlet (pipe); see Figure 14-8L. This should be considered carefully in distributor design, and the volume discharge per square foot of flat tower crosssection must be uniform as possible. Careful layouts of nozzle arrangements are usually required, (Figure 14-8K). Maximum rate variation of ±5–6% of average flow is necessary in design [82].

There are many other types and variations in addition to those listed, although they are usually special-purpose trays and not necessarily generally adaptable.

Good design generally considers that the streams of liquid should enter onto the top of the packing on 3-to 6-in. square centers for small towers (less than 36-in. in diameter), and should number $(D/6)^2$ streams for 36-in. and larger towers, where D is the tower inside diameter, in inches [22]. When the liquid stream spacings exceed 6-in. square pitch, consideration should lean to this figure. Most manufacturers make some type of distributor giving one stream for every 6-in.^{2} of tower area.

The number of irrigation or “drip-points” or entrance points per square foot of flat surface of the tower should be uniform for orifice, weir-type gravity, or pressure distributors, and need not exceed 10 points/ft^2 [82]. This uniformity must not be disturbed by support rings

that support the distributor itself. The distribution must include area adjacent to the wall, and the design must not “force” more liquid at the wall where it contacts the packing. Uniformity of points of distribution to the packing surface is extremely important. The volume flow per point must be carefully calculated.

Bonilla [131] presents an excellent examination of liquid distributors in packed towers. Those with random or structured packing are more sensitive to poor or non-uniform distribution of liquid than tray towers. This requires that liquid and vapor are evenly distributed when they enter the packing. Often, only the liquid distributor at the top of the packing is considered, but vapor distribution at the bottom or middle of the tower is important. The ultimate performance of a packing depends significantly on the initial distribution [131], with non-uniform distribution resulting in reduced packing efficiency, which can be expressed as a higher height equivalent to a theoretical plate/stage/tray (HETP). A structured packing with higher surface area is more sensitive to initial non-uniform distribution of liquid than a lower surface area packing.

It is important to recognize that each packing has a natural liquid distribution [131] that will develop if sufficient bed depth is available. If the distribution is poorer than the natural distribution, the system will end up with concentration gradients and higher HETP values. When/if an improvement from a poor to natural distribution occurs slowly over many feet of packing, Bonilla [131] states that adding extra packing to a bed to compensate for the initial maldistribution does not work, since the return to the natural distribution is not fast enough to compensate for the concentration gradients that have already formed.

Most liquid distributors feed onto the packing by gravity, rather than being pressurized. For any given tower design, the distributor design

and installation are important factors for aiding the packing to do its job. For best performance for the average distillation, the distributor should be installed horizontally to a tolerance of ±⅛ to in. and be leveled once it is in the tower. Table 14-18 [131] presents a comparison of many of the factors necessary to the selection and design of types and styles of distributors. Selection of types depends on liquid rate, turndown requirements, plugging or fouling tendencies, feed flow and many other considerations [131]. The abbreviations in Table 14-18 are for four basic types of distributors described in the following sections:

Table 14-18 Guidelines for Selection of Liquid Distributors

Reproduced by permission of The American Institute of Chemical Engineers; Bonilla, J. A., *Chem Eng. Prog.* Vol. 89, No. 3 (1993) © p. 47; all rights reserved.

Factor	Type of distributor					
	PAN (RTD)	**POH (gravity)**	**POH (pressure)**	**NTD**	**SNH**	**VND**
Uniformity	VG	VG	F	VG	P	P
Solids handling	P	P	F	G	G	VG
Turndown	G	G	G	G	P	P
Ease of installation	F	F	G	F	VG	G
Ease of leveling	F	G	VG	F	VG	F
As redistributor	G[1 and 2]	No[2]	No	No[2]	No	No[2]
Height requirement	M	L	VL	H	M	H
Cost	H	M	L	H	L	M
Residence time	H	L	L	M	L	M
Suitable for large diameters (>10 ft)	P	G	G	G	VG	G
Leakage potential	H[3]	No	No	No	No	No
At high vapor rates	P	G	G	G	VG	G
At high liquid rates	G	G	G	G	VG	G

At low liquid rates	G	VG	G	VG	P	P
For high-purity fractionation	VG	VG	P	VG	P	P
Heat transfer	G	G	F	G	VG	G
Liquid feed handing	Yes[1]	No	No	Yes	No	Yes
Flashing feed handling	Yes[1 and 2]	No[2]	No	No[2]	No	No[2]
Key: VG = very good; G = good; F = fair; P = poor; H = high; M = moderate; L = low; and VL = very low. "No" means that the distributor cannot be used for the particular factor.						
[1]No significant liquid mixing provided, however.						
[2]Very good if used in conjunction with a chimney tray.						
[3]Leakfree if seal-welded.						

14.2.3. Pan Distributors (PAN)

Also known as riser tube distributors (RTD) [131]; these consist of a flat tray with orifices uniformly spaced in order to allow liquid to flow onto the packing below. This tray has vapor risers, uniformly spaced, but not interfering with the liquid orifice holes. They are used for clean liquids in towers of 2–10 ft diameter [131]. They can be installed between flanges in towers smaller than 2 ft. It is good for low to high liquid rates to the tray. With 2-¼ in. diameter risers, about 20 uniformly spaced drip points/ft^2 can be achieved, and this is useful for high purity fractionation. When flows of vapor and liquid are high, the separation efficiency of the system may not be adequate when only 4 to 6 drip points/ft^2 are installed, due to the space occupied on the tray by the larger vapor risers. For high vapor flows, the design must be carefully evaluated [131]. For uniform hole performance, all holes must be punched in the same direction, i.e., top down or bottom up.

Equation 14-1 is used to correlate the number of holes to the liquid head for a specific hole diameter, d:

(14-1)

(14-2)

(14-3)

Symbols are used by permission of Bonilla, [131]:

A	=	distributor quality factor
B	=	distributor quality factor
C_f	=	vapor capacity factor
	=	[(vapor velocity) (vapor density)/(liquid – vapor densities)$^{0.5}$]
DQ	=	distributor quality
a	=	surface area of an orifice hole, in.^{2}
d	=	diameter of hole, in.
h	=	liquid height over orifice, in.
h_d	=	vapor pressure drop across distributor, in. (Calculate h_d by the vapor flow equation through distributor open area)
n	=	number of holes
Q	=	liquid volumetric flow rate, gpm
K	=	flow coefficient
V	=	horizontal liquid velocity, ft/sec (in distributor)

This is considered to be the base design for the distributor [131]. Average values of K are around 0.16, although this is strongly influenced by the direction that two holes are punched in; K is around 0.6 if holes are punched downwards [131].

For distributors of any given design, including the PAN, it is important to filter the feed or reflux liquid entering the distributor, in order to reduce the possibility of plugging the orifice holes. Otherwise, random plugging will cause non-uniform distribution of liquid onto the packing below. It is important to avoid leakage around the risers because this can destroy the liquid distribution pattern [131].

The use of narrow trough PAN distributors is a better choice to prevent leakage from towers above 3-9 ft diameter and larger [131].

Turndown on a PAN type tray should be limited to 2:1 (ratio of high to low flow rates), which results in a reasonable design. Bonilla [131] points out that it is costly to design for short-term high turndown rates, such as startup, shutdown. Increased internal loads are better achieved by increasing reflux ratio during such period, rather than designing the distributor for large turndown.

14.2.4. Trough Distributors (NTD)

These consist of multiple troughs 3–4 in. wide [131], and are fed by feed or parting boxes mounted above, or by multiple pipes mounted at 90° to the direction of the troughs. The parting boxes distribute the liquid to troughs via calibrated orifices in the bottom or wall. The feed points out of the troughs to the packing are usually in the bottom or at the wall of the troughs. Locating the holes in the trough sidewalls allows for collection of rust and any other sediment in the trough bottoms and avoids plugging the orifice holes. This style can handle large vapor loads and the distributor design allows for good liquid distribution. These troughs can be readily leveled across a tower, can still handle low to medium liquid rates, and can handle turndown of up to 5:1 by special and careful design [131]. They should not be used for slurry systems; rather, a V-notch version is better suited, but may not be as accurate for liquid distribution.

Trough distributors are quite often used with structured packing, where it may be desired to have an overflow of a sheet of liquid onto the packing and (rather than through holes) oriented at 90° to the top structured layer [131].

With a trough distributor it is often more difficult to obtain uniform distribution next to the tower wall than with a PAN or orifice pipe distributor. Bonilla [131] recommends the following rule of thumb: Any 10% of the outer surface area of packing in the tower should not receive more or less liquid than the average 10% of the surface area.

The use of V-notches in a trough wall for overflow is more sensitive to leveling problems than other designs, and for the same ±⅛–-in. tolerance produces a more severely non-uniform flow distribution. The quality of distribution from a V-notch is poor compared to the other types of trough distributor, but does have advantages in slurry systems [131]. It should not be used for critical distillation applications, but is good for heat transfer, and where solids present are in the system.

14.2.5. Pipe Orifice Headers (POH)

These distributors have pipe lengths tied to a central distribution header (usually), with orifices drilled in the bottom of the various pipe laterals off the header. This style of distributor can be fed by pressure or by gravity for clean fluids. Gravity feed is considered better for critical distillation applications, when uniformity of the flow from the drip points (or flow points) throughout the crosssection of the tower is extremely important. Gravity is excellent for low flow requirements such as below 10 gpm/ft^2 [131].

This design is restricted to a 2:1 turndown and is not practical for large liquid flow rates compared to the NTD or RTD styles [131].

14.2.6. Spray Nozzle Headers (SNH)

These are similar in design to the pipe orifice distributor using small angle (<90°) spray nozzles instead of orifices. Because the sprays can be selected to cover varying crosssections per spray, the total number can be small compared to the orifice holes that would be required. The spray from the nozzle should be “full cone” and not “hollow cone” to provide a uniform liquid circle. The circles should overlap to avoid dry-spots. The spray nozzles operate under pressure, with the manufacturer providing flow rates and patterns for variations in system pressure. A careful layout is required to control how much liquid flows through each square foot of tower and therefore the packing. This style of distributor is good for heat transfer and vapor washing, with little or no fractionation, although they have been used successfully for distillation operations.

Turndown is usually limited to 0.5:1, and liquid distribution can be poor if the sprays are not carefully laid out, and the system flow tested for uniformity. Another problem is misting of the liquid from the sprays, and the resulting entrainment of liquid out of the tower or up to overhead mist eliminators.

14.3. Number of Flow or Drip Points Required [131]

1. For low purity hydrocarbon fraction, the number of drip or separate flow points should be 6–10 drip points/ft^2 of tower crosssection surface area.
2. For high purity hydrocarbon fractionation, the number of drip points that is recommended for good irrigation is 10–20 drip points/ft^2. Increasing the number of drip points much beyond 20 may not help in improving packing performance. Table 14-19 suggests the approximate number of useful drip per square foot points for various sizes of random packing. Experience and design judgment may indicate when more or less should be used. From Table 14-19, Bonilla points out that at lower flow rates into the distributor, a larger number of holes is required to wet the

packing effectively, due to the lower spreading rates from the lower flows.

Table 14-19 Suggested Number of Drip Points/Ft^2 for Random Packings.

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Regular fractionation	**High liquid loads**	**Low liquid loads**
Small packings	8	10
Midsize packings	6	8
Large packings	6	6
High purity fractionation	**High liquid rates**	**Low liquid rates**
Small packings	12	14
Midsize packings	10	12
Large packings	10	10
Note:		
Small packings include Pall rings 1 in. and smaller, CMR No. 2 and smaller, IMTP 40 or smaller, and Nutter ring No. 1 or smaller; structured packings with ¼ in. crimp and smaller.		
Medium packings include Pall rings 1-½ in. to 2 in., CMR No. 2.5 to 3, IMTP 50, and Nutter ring No. 1.5 and 2.0; structured packings with ⅓ in. to ½ in. crimps.		
Large packings include Pall rings 2-½ in. and larger, CMR No. 4 and 5 and larger, IMTP 70 or larger, and Nutter ring No. 2.5 or larger; structured packings with crimps larger than ½ in.		

3. Minimum recommended hole diameters [131] are:
    a. Carbon steel system: ⅜-in. (use filter on entering liquid)
    b. Stainless steel: ⅛-in. (experience suggests [illegible]-in. minimum) (use filter on entering liquid)

4. Recommended [131] rule of thumb to avoid liquid entrainment

from the entering liquid by the vapor out of the tower: Limit the vapor capacity factor C_f to about 0.4 ft/sec,

where

(14-4)

A value of C_f greater than 0.8 ft/sec in the distributor openings is likely to flood the distributor, or result in heavy entrainment.

Uniform distribution is essential for efficient use and performance of the packing. The distribution of drip points onto the packing should ensure the same amount of liquid wets the packing over each square foot of tower crosssectional area. The area near the wall must also receive the same share of the liquid by ensuring that the number of drip points per square foot is the same right up to the vessel wall i.e., to within one pitch of the orifice holes from the trough or PAN distributor. Bonilla [131] suggests the following rule of thumb for distributor design: The outer 10% of the surface area of the tower cross section should have 10% of the total liquid rate; this means 10% of the number of holes. For some situations, a separate liquid line leading from a point on the distributor to a void in the layout area may be necessary to move liquid to that location. Despite earlier suggestions, Bonilla [131] emphasizes that the vessel wall should receive the same liquid unit quantity as the rest of the packing. The general performance on separation HETP for poor liquid flow in the wall area of the packing is lower compared to the other regions of the column cross section.

For flashing feeds entering between packed sections, it is best to use a chimney type tray, which allow the liquid to mix, and then allow this

to flow onto a NTD or PDH style tray.

For orifice trays, a minimum of 2 in. of liquid should be held on the tray. This varies with size and shape of the drain holes. Figure 14-9 illustrates the effects of different base liquid heights over the hole, and the resulting variation in liquid distribution for a tray or trough that is ¼-in. away from horizontal, comparing the round and slot orifice, and the V-notch.

Figure 14-9

Liquid maldistribution is a function of opening, shape, and liquid head over the opening for a ¼-in. out-of-level distributor. The openings were ¼-in. dia. orifice, ¼-in. slot, and 22.5° V-notch.

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Parting boxes and troughs are so sensitive to the system flows that after the engineer has established these, the services of a competent manufacturer should be consulted. Therefore, due to the detail required, design calculations are not presented in this text.

The downflowing liquid from an upper to a lower packed bed is re-collected onto a collection and redistribution tray. Flow from this tray should be treated as a distribution problem similar to the entry of the liquid onto the top bed. Generally, the height of a bed of packing before collection of downflowing liquid varies with the packing size, style, and the total bed height. As a guide, redistribution should occur

after 10 to 20 ft of packing for random style; but it is advisable to follow the recommendations of the manufacturer.

Overall, careful design of distributors at the top of towers and redistribution of liquid in multi-section towers significantly improves the performance of packed towers. However, the liquid flow is the only important factors; the uniformity of vapor distribution is also important, because non-uniform vapor distribution can cause non-uniform liquid downflow. Selection of packing with the necessary characteristics and requirements/sensitivity to the uniform distribution of the liquid and vapor is another element in process design.

As earlier emphasized, the level of the distributor tray or trough can also be critical to the consistent uniform liquid distribution.

For stacked packing, the liquid usually has little tendency to cross-distribute, and thus moves down the tower over the same crosssectional area that it enters. If packing is constructed by dumping, most follow a conical distribution down the tower, with the apex of the cone at the liquid impingement point. After about 12 ft vertical height, the liquid flows vertically downward unless redistributed. For uniform liquid flow, and reduced channeling of gas and liquid to maximize efficiency, the impingement of the liquid onto the bed must be as uniform as possible.

The liquid tends to flow to the wall, and any reaching it is reduced in effective contact possibilities, so no more than 10% of the total liquid should enter the packing at the top packing-wall circumference, or within the 5–10% of the tower diameter next to the tower wall. With this approach, the bulk of the liquid starts to flow down the tower somewhat away from the wall.

When using plate-type distributors, an out-of-level condition can

cause serious channeling of liquid down one part of the column and gas up the other. Provision such as V-notches, should be made for self-adjusting levels of liquid, which will allow for any shifting of tower alignment, brick walls, *etc.* Gas velocities through the tower at the point of leaving the packing, and/or through the distributor plate gas risers should be low, in order to reduce liquid carry-through. This can be calculated by using liquid entrainment limitations. From limited tests, it appears that there is essentially no entrainment off a packing until the flooding point is reached.

The various packings have characteristic ways of distributing the liquid throughout the bed. Leva [40] shows the results of Baker *et al.* [3], which illustrate the effect of various types of distribution on the liquid pattern inside the packing. A general summary is given in Table 14-20.

Table 14-20 Liquid Distribution Patterns in Packed Columns (Data in 6-in., 12-in., and 24-in. Dia. Towers).

*Compiled from M. Leva. *Tower Packings and Packed Tower Design*, 2nd Ed., U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.) (1953), Ref. 40, by permission.

Packing	**Tower dia. in.**	**Type liquid feed**	**Percent liquid distribution in inner ¾ tower area**	
			2 ft from top	**8 ft from top**
½-in. Raschig Rings	6	Center Point	40	40
	12	Center Point	90	78 (4 ft down)
	12	4 Point	70	65 (4 ft down)
1-in. Raschig Rings	2	Center Point	95	70 (4 ft down)
	2	4 Point	70	60 (4 ft down)

½-in. Berl Saddles	6	Center Point	60	50
	12	Center Point	95	88 (4 ft down)
	12	4 Point	70	60 (4 ft down)
1-in. Berl Saddles	12	Center Point	85	65 (4 ft down)
	12	4 Point	75	55 (4 ft down)
1-in. Lessing Rings	12	Center Point	90	55
	12	4 Point	70	55
	12	19 Point	70	50
	24	Center Point	100	90
	24	12 Point	80	70
½-in. Glass Rings	6	Center Point	90	70 (7 ft down)
	6	Center Point	83	75 (7 ft down) 3 ft/sec, air
	6	4 Point	75	70 (7 ft down)
	6	4 Point	75	75 (7 ft down) 3 ft/sec, air
½-in. Spheres	6	Center Point	83	85 (4 ft down)
	6	Center Point	90	90 (4 ft down, 2.5 ft/sec, air

Strigle [82] has established that the necessity for uniformity of liquid rate increases as the number of theoretical stages per packed bed increases. Below five theoretical stages the packing is not so sensitive to the uniformity of liquid distribution. With more than five stages per bed, the liquid distribution has a significant effect on the packing efficiency. Larger packing sizes are less sensitive to the uniformity of liquid distribution than the smaller sizes.

Norton [83] classifies liquid distributors as:

1. High performance Intalox®: 90–100% quality
2. Intermediate performance: 75–90% quality
3. Historic standard: generally 30–65%.

The quality grading system relates to uniformity of distribution across the tower crosssection, where 100% quality indicates fully uniform distribution. Distributors are designed to suit the system, particularly the packing type and size.

The type of distribution needed depends on the sensitivity of the tower performance to the liquid distribution as discussed earlier. Norton's [83] data indicate that the sensitivity of tower performance to liquid distribution depends only on the number of theoretical stages in each bed of packing that is achievable at its “System Base HETP” [83]. Tower beds of high-efficiency packing are more sensitive to liquid distribution quality than shorter beds of medium-efficiency packing [83]. It is important to extend the uniformity of distribution to within one packing particle diameter of the tower wall [85].

Good liquid distribution starting from the distributor and at the top layer of packing is essential to fully utilize the bed [85]. In principle, this applies to all types and sizes of packing. Kunesh *et al.* [84, 85] present Fractionation Research, Inc. (FRI) studies on distribution that reflect the importance of maintaining level distributor trays (devices), and of eliminating discontinuities or zonal flow (Figure 14-10). Their results further show that a “packed bed has a reasonable tolerance for both a uniform or smooth variation in liquid distribution and for one that is totally random”. Once a system has become maldistributed, regaining natural and constant HETP is very difficult.

Figure 14-10

Effect of liquid maldistribution on efficiency; FRI data for 25-mm Pall rings in cyclohexane/n-heptane distillation with two different quality distributors.

Used by permission of the American Institute of Chemical Engineers, *Chemical Engineering Progress*, Perry, D. and Nutter, D., Jan. (1990) p. 30, and by special permission of Fractionation Research, Inc., all rights reserved.

Perry *et al.* [85] point out that packed columns are more dependent on liquid distribution than trayed columns, as can be appreciated by the differences in the way the liquid must flow down the two types of column. Liquid distribution quality is measured or described as [85]:

1. Number of distribution points (distribution density).
2. Geometric uniformity of distribution points across the crosssection of the tower.
3. Uniformity of liquid flow from the distributor points.

Currently, most designs use 4 to 10 distribution points per square foot of tower crosssection, with 9 points being generally considered optimal for a wide variety of random and structured packings [85]. The distribution demands of small random packings are greater than for the larger sizes due to the lower radial spreading coefficients, i.e., the larger the radial spreading coefficient the more quickly the initial liquid distribution will reach an equilibrium with its “normal” or “natural” frequency (see Figure 14-11).

Figure 14-11

Radial spreading coefficients for several types and sizes of packings. Small packings require better initial distribution. The larger the radial spreading coefficient, the more rapid the initial distribution will reach its natural equilibrium of flow distribution. After Hoek, P. J., Wesselingh, J. A., and Zuiderweg, [86], and Nutter, D. [88]; reproduced with permission of the American Institute of Chemical Engineers; Perry, D., Nutter, D. E., and Hale, A., *Chemical Engineering Progress*, Jan. (1990), p. 30; all rights reserved.

Hoek [86] proposed a radial spreading coefficient to characterize liquid distribution. This coefficient is a measure of how quickly a packing can spread a vertical liquid stream radially, as the liquid flows down the column [86]. Radial mixing tends to reduce the effects of poor efficiency due to concentration gradients brought about by liquid maldistribution. The “spreading coefficients” of structured packing are larger than for random packing. The spreading for structured packing occurs in a somewhat different manner due to the layering of the structured packing.

The entrance of a liquid-flashing vapor mixture into the distillation column at the feed location requires a specially designed distribution tray in order to separate vapor from liquid, which must drop onto the packing bed for that section in a uniform pattern and rate.

Structured packing requires specially designed distributors, as recommended by the respective manufacturers to ensure uniform liquid distribution across any bed of this type of packing.

14.4. Redistributors

The liquid coming down through the packing and on the wall of the tower should be redistributed after a bed of depth of approximately 3 tower diameters for Raschig rings, and 5–10 tower diameters for saddle packings. As a guide, Raschig rings usually have a maximum of 10–15 ft of packing per section, while saddle packing can use 12–

20 ft. This redistribution brings the liquid off the wall and outer portions of the tower, and directs it toward the center area of the tower. This ensures uniform distribution and effective contact in the lower section.

The redistributor must be sealed against the tower wall in order to collect all of the liquid from the packed section above (Figure 14-11). It must be capable, either singly or in conjunction with a distributor placed below it, of redistributing the collected liquid in an efficient manner. (See discussion in previous paragraph on distributors.) The gas/vapor riser opening must be designed to avoid liquid dropping directly through the gas risers and onto the packed section below. This vapor flow area must be relatively large in order to avoid localized development of high pressure drop and subsequent are in performance.

The height of packing before redistribution is a function of the liquid flow pattern through the packing, and this inturn depends on the size and type of packing. Some towers can have 20–30 ft of packing with no redistribution; however, this may be determining by economic as well as operational factors. The exact amount of performance efficiency sacrificed is questionable, although with 20–35% of the liquid flowing down the walls after 10 ft of ring packing depth, it appears that contact is lost for most of this liquid. Redistribution is usually not necessary for stacked bed packing, as the liquid flows in essentially vertical streams. However, most packed tower services do not use stacked packing.

Physically, the redistributors may be a simple and relatively inefficient side wiper as in Figure 14-12 and Figure 14-13; a conventional support grid, or plate plus regular distribution plate as used at the top; a combination unit similar to Prym support and distributor; or a support plate as shown in Figure 14-14Figure 14-7D

and Figure 14-7E a circular plate with holes. The possibility of causing flooding in the tower at the redistribution point must not be overlooked, as too much restriction by a wall wiper, or by packing on a plate can be the focal point for poor tower performance. The velocity conditions should be checked for the smallest crosssection.

Figure 14-12

Metal rosette style redistributor.

Used by permission of Norton Chemical Process Products Corp., Bull. TA-80.

Figure 14-13

Typical metal full crosssection redistributor.

Used by permission of Norton Chemical Process Products Corp., Bull. TA-80.

Figure 14-14

Liquid redistribution in packed towers.

Used by permission of Norton Chemical Process Products Corp.

14.5. Wall Wipers or Side Wipers

The wall wiper liquid collector/redistributor, Figure 14-12, is most useful in reducing the bypassing effects of liquid running down the walls of small towers. They do not properly replace redistributor systems placed periodically in the tower. For larger towers of 4 ft diameter and greater, they are not as useful, since they collect a smaller portion of the total tower liquid, and cannot effectively redistribute it throughout the tower crosssection– as discussed in the paragraphs under “Distributors”. They are useful for smaller towers, of about 18 in.–20 in.–30 in diameter. They can restrict vapor flow up, as well as inhibit redistributing the collected liquid uniformly across the tower.

14.6. Hold-down Grids

To reduce ceramic or carbon packing breakage and blowing out of light weight plastic packing when a tower surges due to gas pockets, uneven loading, etc., it is sometimes helpful to have heavy hardware cloth or other type of stiff but open grid resting (floating) at the top of the tower, and on the top of the packing (Figure 14-15A and Figure 14-15B).

Figure 14-15A Hold-down plate for packed tower.

Used by permission of US. Stoneware Co. (now, Norton Chemical Process Products Corp.).

Figure 14-15B

Typical metal hold-down plate for use with ceramic or carbon packing. Note: It rests directly on top of the packing; bed-limiters are similar in design in metal or plastic, bolted to column wall above packing.

Used by permission; Norton Chemical Process Products Corp., Bull. TA-80.

This grating or grid must be heavy enough to hold down the top five feet of packing, yet must be able to move down as the packing settles, always resting at its top. If the packing is restricted in upward movement, it usually will not be crushed but if it does break and crush, the bed settles and its characteristics change considerably.

Bed-limiters are usually light-weight and must be bolted in place, not resting on the packing. They are used with metal and plastic packings to prevent the bed lifting, or individual pieces of packing being entrained, and carried out of the tower. These packings usually do not break, and as long as the bed temperature is below the softening or deflection point for plastic packing, the bed should not compress or consolidate to create operational limitations.

14.7. Packing Installation

Stacked

Preparing stacked packing is a hand operation and rather costly. It is avoided where possible except for the initial layers on supports. Liquid distributed on a stacked packing usually flows straight down through the packing that is immediately adjacent to the point of contact. There is very little horizontal liquid flow. Packing patterns perform differently, and are illustrated in Figures 14-16A–C.

Figure 14-16

(A) Stacked packing: square pattern (S). (B) Stacked packing: diamond pattern (D). (C) Stacked packing: packing assembly.

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

Dumped

Dumping is the most common method of installing packing. If possible, the tower should be filled with water (for ceramic packings) after installation of the bottom support arrangement, and the loose packing floated down to rest on top of the support. The fall should be as gentle as possible, since broken packing tightens the bed and increases pressure drop.

One manufacturer [22] suggests loading ceramic ring type packing by using a sheet metal cone in the tower, thereby causing the rings to slide off the cone and fill to the edges first. The cone is lifted as the rings are floated or dropped in. This technique is believed to reduce the tendency for rings to channel liquid to the tower wall. Saddle-type packing does not require the use of the cone.

In packing a tower dry, high hydrostatic heads are thus avoided on joint connections; however, extra care must be taken to avoid ceramic breakage. The packing will probably settle after installation, but it should not be pressed or tamped in to place, as this will cause extra breakage. Further, the packing should not be allowed to fall more than two feet to the bed surface; it should be dumped at random to avoid any pattern developing. The dry packed tower will be denser than the wet packed. The pressure drop for dry packed beds can be as much as 50–60% greater than for wet packed. Dry packing is not the preferred method due to significant settling that will occur.

Metal and plastic packing can be dumped into the tower when dry; however, reasonable care must be exercised, and the drop should not

be more than 10–15 ft, using a chute with a cloth sock (open) on the end to cushion the impact.

Repacking a tower will usually change the pressure drop; for small 8-in. diameter units this may be as much as 100%. In larger columns (24-in. diameter and higher), this variation is appreciable; about 50% or less.

14.7.1. Packing Selection and Performance

Guide Lines: Trays versus Packings

Kister *et al.* [136] prepared one of the few comprehensive distillation studies for the selection of valve and sieve trays compared to random or structured packing. This reference is based on a more comprehensive evaluation of accumulated data by the same authors [137]. Many separate studies have been conducted for trays [138] including bubble caps as well as various packings, but few, if any, attempt to establish similar conditions in order to make a viable comparison as is attempted in References 136 and 137. There are two main considerations related to capacity and separation [136].

1. An optimal tray design is one that balances tray and downcomer area so that neither prematurely restricts capacity. Weir height, weir geometry, clearance under the downcomer, and fractional hole area are set to maximize efficiency and capacity.
2. Optimal packing design configures distributors, supports and bed heights so as to avoid loss of efficiency to maldistribution and premature capacity restriction [136].

The four main differences between bubble cups and packings are [136]:

1. Differences between the capacity and efficiency of an optimal tray and an optimal packed tower design.
2. Deviations from optimal design of trays, packings and other tower internals, e.g., distributors and baffles.

3. Unique system characteristics and special design features, e.g., corrosion, foaming, chemical reaction, and fouling, and designs to overcome such problems.
4. Capacity and separation gains due to lower pressure drop of packing. Pressure drop of packing is typically 3 to 5 times lower than that of trays.

Due to the need to use case-by-case analysis, the Kister studies [136, 137] focused on item 1. The data evaluated came from published reports by Fractionation Research Inc. (FRI) and the Separation Research Program (SRP) at the University of Texas, taken from commercial size equipment rather than laboratory research columns. The FRI data include No. 2 and No. 2.5 Nutter random rings packing, and Norton's Intalox® 2T structured packing. Each considered currently state-of-the-art or close to it, while the sieve and valve trays were of FRI's latest designs, plus Nutter's proprietary valve trays, all using 24-in. tray spacing.

To allow for the vertical height required in packed columns for tower distributors and redistributors–typically 10%–20% of the vertical packed height (10% for 2-in. random packing and 20% for structured packing) [136] and in tray towers, the height required by additional trays, the analysis indicated:

(14-5)

For practical trays HETP,

(14-5A)

where

m = factor higher than test HETP; = 1.1 for 2-in. random packing
= 1.2 for structured packing used
S = tray spacing, in.
E = overall column efficiency, %

Capacity factor:

(14-6)

Tray spacing [136]:

(14-7)

Correlating liquid rate and pressure, see Figure 14-17.

(14-8)

(14-8A)

where

C_S	= capacity factor, ft/sec, based on tower superficial area
E	= overall column efficiency, trayed column, %
FP	= flow parameter, dimensionless
HETP	= height equivalent of a theoretical plate, in.
L	= liquid flow rate, lb/hr-ft^2 of crosssectional area

m	= constant, allowing for vertical tower height consumed by distribution/redistribution equipment
S	= tray spacing, in.
U	= velocity, ft/sec based on tower superficial area
U_G	= superficial velocity based on crosssection area of empty column, ft/sec
V	= vapor flow rate, lb/hr-ft^2 of crosssectional area
ρ	= density, lb/ft^3
Fp	= packing factor, empirical
ν	= kinematic viscosity, liquid, centistokes (kinematic viscosity = viscosity, centipoise/specific gravity) (not density)

Figure 14-17

Overall comparison of capacity at flood for 24-in. tray spacing with random packing.

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Subscripts: G = gas; L = liquid

Figure 14-17 plots flood capacity versus flow parameter. The FP values of 0.4−0.7 have been estimated by Kister *et al.* [136] in absence of data. The plots show that for low and moderate pressures, the flood capacity factor versus FP correlates the effects of liquid rate and pressure on the optimized tray capacity [136]. At higher pressures an additional effect of pressure on capacity shows a decline of optimized tray capacity.

Figure 14-18 plots HETP versus FP for optimized trays at 24-in. spacing, and No. 2 and No. 2.5 Nutter Rings and Intalox® 2T structured packing.

Figure 14-18

Overall comparison of efficiency for "state-of-the-art" random and structured packing with trays at 24-in. spacing.

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In summary, we quote the conclusions of Kister *et al.* [136]:

"Comparing trays at 24-in. tray spacing with a state-of-the-art 2−2.5-in. (nominal) random packing, and with a state-of-the-art structured packing of 67 ft^2/ft^3 specific surface area, all optimally designed, we found that:

At FPs of ≅ 0.02 to ≅ 0.1:

- The trays and the random packing have much the same efficiency and capacity.
- The structured packing efficiency is about 50% higher than either the trays or the random packing.
- As FP increases from 0.02 to 0.1, the capacity advantage of the structured packing (over the trays or over the random packing) declines from 30–40% down to 0.

At FPs of 0.1 to 0.3:

- The trays and the random packing have much the same efficiency and capacity.
- The structured packing has much the same capacity as the trays and the random packing.
- As FP increases from 0.1 to 0.3, the efficiency advantage of the structured packing over the random packing and over the trays declines to about 20% from about 50%.

At FPs of 0.3 to 0.5:

- Efficiency and capacity for the trays, the random packing, and the structured packing decline with a rise in flow parameter.
- The capacity and efficiency decline is steepest in the structured packing and shallowest in the random packing.
- At an FP of 0.5 and 400 psia, the random packing appears to have the highest capacity and efficiency and the structured packing the least.

The above results are based on data obtained for optimized designs and under ideal test conditions. To translate our findings to the real world, one must factor in liquid and vapor maldistribution, which is far more detrimental to the efficiency of packings than of trays. In addition one also must account for poor optimization or restrictive internals, which are far more detrimental to the capacity of trays than packings".

Chen [133] highlights the long-term growth of the use of bubble cap trays, valve and sieve trays, followed by the increased popularity of packed columns accompanied by the development of random and structured packings. There are some applications in chemical/petrochemical/petroleum/gas treating processes where one type of contacting device performs better and is more economical than

others. Chen [133] points out:

1. A typical tray has an opening area ranging 8% to 15% of the tower crosssectional area.
2. A typical packed tower design has more than 50% of open tower crosssection, with the void fraction of a packed tower being higher at around 90% of tower volume, resulting in the following:
    a. **Pressure drop per theoretical stage** − Packed towers usually result in lower pressure drop per theoretical stage than trays. Trays often show drops of 3 to 8 mm Hg per theoretical stage, with packing having about 1 to 2 mm Hg for random packing and 0.01 to 0.8 mm Hg for structured packing. For high pressure systems, the difference may not be significant, but for atmospheric and below atmospheric pressures, but it can be quite significant.
    b. **Liquid hold-up** − Trays usually hold-up 8 to 12% of tower volume, compared to 1 to 6% for packed towers. This can be significant for systems involving thermal degradation and that need very short residence times, which also aids in sharp separations.
    c. **Liquid/vapor ratios** − Trays are designed for low liquid/vapor ratios, while packed towers are operated from low to high liquid/vapor ratios (often in absorbers and scrubbers). Low L/V ratios are usually associated with distillation, with (approx.) less than 10 gallons of process liquid/(min) (ft^2) (tower crosssection area).
    d. **Liquid cooling** − This is usually better handled in tray towers, and it is easier to draw off liquid from trays for removal from the system, or for external recirculation.
    e. **Foaming systems** − The contact surfaces of packings promote film action compared to the droplets from trays for mass and heat transfer: packings tend to be more resistant to entrainment and induce less foaming.

f. **Corrosion** − Corrosion problems with some fluid systems dealt with more easily and at lower cost by corrosion resistant packings than by fabricated trays.

g. **Solids and slurries** − Trays will handle solids and slurry systems better than packings; and if solids buildup does occur washing/flushing treatment will usually “wash” or dissolve the solids attached to the trays more easily than by attempting a thorough cleaning of packing.

h. **Costs** − Other than special needs requiring one contacting mechanism or another, small-diameter columns of 18-in, diameter or less can be assembled less expensively as packed towers. For some applications of larger diameter columns, the packed tower may still provide the less expensive choice. A cost analysis comparison would determine benefits for individual situations.

Chen [133] recommends the following guidelines for the design of the important distributors of liquid (still must pass vapor):

Pan-type distributor: Plate with drilled/punched holes for liquid downflow and vapor risers.

vapor riser: 15 to 45% of tower crosssection area, round risers usually 4 in. or 6 in. diameter, although the round design usually has less free area than a rectangular design. Usual standard height is 6 in., however, any height can be used as long as it is well above the liquid height on the tray. The pressure drop through the vapor riser should be low:

(14-9)

Liquid orifices on tray pan: usually at least 10 orifices per square foot of tower area. The orifice diameter can be determined [133] by:

(14-10)

where

ΔP = pressure drop through risers, in. of liquid
V = vapor rate, ft^3/second
A = riser area. ft^2
D_v = vapor density, lb/ft^3
D_l = liquid density, lb/ft^3
d = orifice diameter, in.
Q = liquid flow rate, gal/min
H = differential head at orifice, ft of liquid
K = discharge coefficient, for punched holes = 0.707

Liquid design height is usually one-half of the riser height. At minimum rates, the depth on the tray can vary from ½ in. to about 1 in. below top of riser height for maximum rates. The minimum orifice diameter is recommended to be ¼ in. diameter to overcome miscellaneous plugging of the holes [133]. Experience indicates the holes really should be ⅜ in. to avoid industrial plugging problems.

Other useful distributor types have been referred to and previously illustrated. For redistribution, the vapor risers may be 12–18 in. tall, and with protective “hats” to prevent liquid dropping from the tray/section above. The space between the cover “hat” on the riser and the bed above should be 18 in. to 12 in. minimum, to allow for proper vapor redistribution entering the packed section above. The importance of a horizontally level distributor cannot be

overemphasized.

Any of the available packings can usually be exchanged with another; the differences being in efficiency of contact, expressed as height of transfer unit (HTU), height equivalent to a theoretical plate/stage (HETP) or Kga, and pressure drop for the particular packing-fluid system. Therefore, system data are very important when selecting a packing. When not available, an effort should be made to find an analogous system in terms of process type, fluids, physical properties, pressure and temperature conditions, *etc.* If this is not possible, then the best judgment of the designer must be used.

Eckert [125] provides some basic guidelines to good packing selection over a range of system performance requirements. Kunesh [126] illustrates the often-determined pressure drop advantage of random packed towers over the usual valve tray. See Figure 14-19 [126].

Figure 14-19

Comparison of typical valve tray and random packing showing that packing reduces pressure drop significantly.

Used by permission of Kunesh, J. G., *Chemical Engineering*, V. 94, No. 18 (1987) p. 101, all rights reserved.

For a preliminary guide and broad comparison of packings versus various distillation tray types, Table 14-21 is helpful. Although the table includes, a list of the more prominent manufacturers of trays and packing materials, it is not all-inclusive. Table 14-22 [123] compares trays, random and structured packing, and HETP, where the C-Factor

Table 14-21 Relative Performance Characteristics of Tower Packing and C

Used by permission, W.P. Stading, *Chemical Processing*®, Feb. (1989), Kirkma

Company	Configuration	Material of const. [2]	Relative cost[3]	Mass transfer[4]	Pressure drop	Capa
GLITSCH, INC.						
Trays						
P-K TraysTM	Trays w/baffles	M	Mod	Med	Lo	V.Hi
Screen Trays®	Venturi effect	M	Mod	Med	Lo	V.Hi
Ballast®	Valve trays	M	Mod	Med	Med	Med
Bubble Cap	Bubble cap	M	Hi	Med	Hi	Med
Random Packing						
Ballast®	Pall rings	M, P	Lo	Med	Med	Med
Cascade Mini-Rings®	Low aspect ratio	M, P, C	Mod	V.Hi	Lo	Hi
Structured Packing						
Gempak®A	Corrugated, perforated, lanced	M, P	Mod	V.Hi	V.Lo	V.Hi
Gempak® AT	Corrugated, perforated, lanced	M	Mod	V.Hi	V.Lo	V.Hi

	lanced					
Glitsch Grid® EF-25A, EF-25AP	Lattice grid	M, P	Lo	Lo	Lo	Hi
York Product						
Structured Packing						
Goodloe®	Knit mesh	M, P	Hi	E.Hi	Hi	Hi
Goodloe®	Bicomponent knit mesh	P + M	Lo	E.Hi	Med	Hi
JAEGER PRODUCTS INC.						
TRI-PACKS®						
(Hacketten® in Germany)	Hollow spherical	P	Hi	E.Hi	Lo	V.Hi
NOR-PAC®	Cylindrical rings/ribs	P	Lo	Hi	E.Lo	V.Hi
METAL VSP®	Hollow slotted ring	M	Mod	E.Hi	Lo	V.Hi
TOP-PAK®	Semi-spherical	M	Lo	V.Hi	Lo	V.Hi
NOVALOX® SADDLES	Smooth, beveled, longitud. ribs	C	Mod	Med	Med	Lo
BERYL® SADDLES	Smooth, beveled	C	Hi	Med	Med	Lo

SADDLES	Slotted longitud. ribs	C	Hi	Med	Med	Lo
Structured Packing						
NOR-PAC® KOMPAKT®	Multi-layers screen cylinder	P	Hi	Hi	Lo	Hi
Metal structural packing	Vertical sheets (New design)	M	Hi	E.Hi	E.Lo	E.Hi
KOCH ENGINEERING						
Trays						
FLEXITRAY® 4 variations	Circular slotted valve	M	Lo	Med	Med	Med
Random Packing						
FLEXIRING®	Slotted cylinder	M	Mod	Hi	Lo	Med
	Internal tongues	P	Lo	Hi	Lo	Med
K-PAC®	Modif					
	FLEXIRING® increased voids	M	Mod	Hi	V.Lo	Hi
FLEXISADDLE®	Saddle	P, C	Lo	Med	Med	Med
Structured Packing						
FLEXIPAC®	Geomtr. arrngd.	M, P	Hi	V.Hi	V.Lo	Hi
(Mellpak® in Europe)	corrug. sheets					
SULZER (KS)		M, P	V.Hi	E.Hi	V.Lo	V.Hi
NORTON						

CHEMICAL PROCESS PRODUCTS						
Trays						
Valve tray	Valve	M	Mod	Med	Med	Med
Sieve tray	Slotted cylinder	M	Mod	Med	Med	Med
Bubblecap	Bubblecap	M	Mod	Med	Hi	Med
Random Packing						
IMTP®	Rib finger type	M	Mod	Hi	Lo	Hi
HY-PAK®	Slotted cylinder	P	Mod	Med	Lo	Hi
SNOWFLAKE™	Short cylinder ribbed	P	Lo	Hi	Lo	Hi
Pall rings	Slotted cylinders	M	Mod	Med	Med	Med
Pall rings	Slotted cylinders	P	Lo	Med	Med	Med
Super INTALOX®	saddle	P	Lo	Med	Med	Med
Saddles						
Super INTALOX®	saddle	C	Mod	Med	Med	Med
Saddles						
Structured Packing						
INTALOX® structured packing	corrugated	M	Hi	Hi	Lo	Hi
INTALOX® wire	woven	M	Hi	E.Hi	E.Lo	Hi
NUTTER ENGINEERING						
Trays						

FLOAT VALVE® tray	Rectangular valve	M	Mod	Med	Med	Med
V-GRID® tray	Tapered slot	M	Mod	Med	Med	Med
Random Packing						
Nutter Rings	Crimped, curved slotted strips	M	Mod/Lo	Hi	V.Lo	Hi
Structured Packing						
SNAP-GRID™	Slotted, snaplock shape	M	Hi	Med	E.Lo	E.Hi
Montz A3™	Wire-weave corrugated	M	V.Hi	V.Hi	Lo	Hi
Montz B1™	Embossed sheet metal	M	Hi	V.Hi	V.Lo	Hi

GENERAL COMMENTS

[1]Comparisons of relative cost and performance are applicable only within same r

[2]M = Metal (Generally 304 SS. Other alloys available); P = Plastic (Wide selecti

[3]Costs: Mod = Moderate or = conventional packing; Lo = conventional packing; packing.

[4]Mass Transfer efficiency.

5. Structured packings frequently used for high vacuum service

5. Structured packings frequently used for high vacuum service.
6. Carbon steel and other metals sometimes available.
7. Wide range of plastics generally available.
8. Packing efficiency and capacity vary with specific application. Contact vendor making final decision.
9. There is no intention to reflect negatively on any manufacturer's packing or tra

Table 14-22 Typical Performance Characteristics Comparison of Tower Packings and Trays

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Characteristic	Type of internal		
	Trays	**Random packing**	**Structured packing**
Capacity			
F-factor, (ft/s) $(lb/ft^3)^{1/2}$	0.25–2.0	0.25–2.4	0.1–3.6
C-factor, ft/s	0.03–0.25	0.03–0.3	0.01–0.45
Pressure drop, mm Hg/theoretical stage	3–8	0.9–1.8	0.01–0.8
Mass transfer efficiency, HETP, in.	24–48	18–60	4–30

Souders-Brown C-Factor:

(14-12)

and F-Factor:

(14-13)

where

V_s = superficial vapor velocity, ft/sec (across tower crosssection)
$D_v = \rho_v$ = vapor density, lb/ft^3
$D_l = \rho_1$ = liquid density, lb/ft^3

Trays are usually designed with an F-factor between 0.25 and 2.0 for a turndown of 8:1. Pressure drop per theoretical stage falls between 3 and 8 mm Hg. Note that bubble cap trays are on the high end, and sieve trays are on the lower end of the range. Varying tray spacing and system efficiency, changes the HETP for trays between 24 in. and 48 in. [133]. The C-factor is given by the familiar Souders and Brown capacity equation.

The number of packing sizes, types (designs), and materials of construction available to the designer has increased considerably. Packing selection for a specific process application requires a weighing of information and an evaluation of the closest comparable data.

Overview

Kunesh [126] presents an overview of the basis for selecting random packing for a column application. To decide between trayed or packed towers–the fundamental design options–a comparative performance design and its mechanical interpretation should be completed, considering pressure drop, capacity limitations, performance efficiencies (HETP), material/heat balances for each alternative. For one example, relating to differences in liquid distribution performance, see Reference 126.

If a packed tower is chosen, the larger packing size generally provides the greater capacity, with less pressure drop, but at the expense of lower efficiency (higher HETP) than a somewhat smaller size. Packing size is the primary variable in determining optimum

performance, with packing size being an important secondary consideration. Other factors, such as column diameter and length devoted to packing, also influence overall performance. Obviously, there is a close balance here, particularly between the various design shapes (types) of the different manufacturers.

For more accurate performance data for a specific packing type/size, consult the respective manufacturers and do not only rely on the generalization charts of the published literature. These charts are continuously being improved, so they are quite useful for a good approximate design (and even final in some instances). Some competitors have packings so close in design to each other that there is little real difference between them, particularly because a reasonable "factor of safety" should be applied more specifically to packing height (when separation of components is more important) than to tower diameter (volume/mass capacity).

Fractionation Research, Inc. (FRI) [126] has found that the following parameters (plus a few others) affect efficiency (HETP): system to be separated, concentration of components, absolute pressure level, column diameter, and bed length, depth, or height (the latter two primarily related to the quality of liquid/gas distribution). Kunesh [126] urges caution when selecting an efficiency prediction (HETP) that is "close to the operating conditions for an accurate/final design". Experience suggests it may be necessary to select a final design HETP from the best available data (family of compounds, pressure of operation and specific packing type and size) and add a factor of "safety" to suit the situation, perhaps 15–30%.

14.8. Contacting Efficiency, Expressed as Kg_a, HTU, HETP

When specific system data are not available, and often they will not be, then close comparisons should be sought. If nothing more can be done, tabulate the relative efficiency for other systems and apply judgment to select a value.

The HETP (Height Equivalent to a Theoretical Plate (stage or plate)) is the tray spacing divided by the fractional overall tray efficiency [82] .The transfer unit concept has been useful for generalized correlations [89]. Because packed towers operate with continuously changing compositions through the packed height, the concept of HETP has been used to determine the number of theoretical stages (plates) required for a given separation by the usual discrete tray-by-tray method (stepwise), and then using the height of packing equivalent to one theoretical plate, and then multiply in to obtain the total height of packing. This requires the use of experimentally, or industrially determined HETP values for the same system (or one quite close in terms of pressure, types, or families of fluids and packing size and family type see Figure 14-18 and Table 14-22 and later discussion). Table 14-23 [133] compares several process systems and gives the corresponding average HETP for 2-in. diameter slotted packing rings.

Table 14-23 HETP Comparison of 2-in. Dia. Slotted Rings

Used by permission: Chen, G. K., *Chem. Eng.* Mar. 5 (1984), p. 48, all rights reserved.

System	Top pressure, mm Hg	Bed depth, ft	Averaged HETP, in.
Acetone/water	740	5	18
		10	20
Iso-octane/toluene	740	5	24
		10	30
		20	32
	100	5	24
		10	26
		20	28
Methanol/water	740	10	21
[illegible]	[illegible]	[illegible]	[illegible]

Isopropanol/water	740	10	24
Para/ortho xylene	100	10	33
	50	10	31
	16	10	28
Ethylbenzene/styrene	100	10	22
		20	26
Basis: Test-column diameter: 24–42 in.			

14.9. Packing Size

This affects contact efficiency; usually, the smaller packing is more efficient; however, pressure drop increases.

As a general guide, use:

Random Packing Size, Nominal, in.	Column diameter in.
½–⅝	6–12
⅝–1	12–18
1–1½	18–24
1½–2	24–48
2–3	36– larger

The packing sizes listed relate to ceramic (where available), metal and plastics. Plastic random packing must be used well under its softening temperature. Due to the packing weight of the bed, the packing can compress as the temperature increases, become less effective, and thus increase the system pressure drop to the point of causing flooding. Therefore, the manufacturer should be consulted regarding upper limits of operating temperatures and corrosion resistance of each plastic. Under these conditions it is important that the plastic packing surface be “wetted” to allow film formation.

14.10. Pressure Drop

The pressure drop, (ΔP), represents frictional and kinetic energy losses and the force exerted by the operating liquid holdup through the packed bed. At constant ΔP, there is less volumetric liquid holdup for high density liquids, but for low density liquids, the volume of liquid holdup can be significantly greater than for water at the same ΔP. At atmospheric pressure and in vacuum service, the static head produced by gas is negligible. In higher pressure fractionators, where more theoretical stages are needed, the static head of vapor from the condenser to the reboiler can be considerable. In such cases, the value should be added to the calculated ΔP in determining the bottom column pressure and reboiler temperature. Pressure drop may be estimated from Figure 14-21C for given packing system properties and operating conditions [159], or may be calculated by the following equation, as represented by Leva [160]:

(14-14)

The equation was originally developed for air–water systems. For other liquids, L^1 is multiplied by the ratio of the density of water to the density of liquid. Table 14-24 and Table 14-24A, show ΔP correlations for different types of packing material.

Table 14-24 Pressure Drop Correlations for Packed Towers

Source: Van Winkle, M., Distillation, McGraw-Hill, Inc. (1967)

Packing		Constants	
Type	**Nominal size,in.**	**α**	**β**
	⅜	4.70	0.41
	½	3.10	0.41
	⅝	2.35	0.26

Raschig rings Ceramic	⅝	2.33	0.26
	¾	1.34	0.26
	1	0.97	0.25
	1¼	0.57	0.23
	1½	0.39	0.23
	2	0.24	0.17
Raschig rings	⅝	1.20	0.28
Metal, 1/32 in.	1	0.42	0.21
Wall, 1/16 in.	1½	0.29	0.20
	2	0.23	0.135
Pall rings, metal	⅝	0.43	0.17
	1	0.15	0.16
	1½	0.08	0.15
	2	0.06	0.12
Berl saddles	½	1.2	0.21
	¾	0.62	0.17
	1	0.39	0.17
	1½	0.21	0.13
Intalox saddles	½	0.82	0.20
	3/4	0.28	0.16
	1	0.31	0.16
	1½	0.14	0.14

Table 14-24A Revised Packed Tower Pressure Drop Correlation Constants for Flooding Region.
By permission J. Eckert, U.S. Stoneware Co. (now, Norton Chemical Process F

Type of packing	**Mat'1**	**Nominal pac**						
			¼	⅜	½	⅝	¾	1
Intalox Saddles	Ceramic	α			1.04		0.52	0.5
		β			0.37		0.25	0.1
Raschig Rings	Ceramic	α			1.96	1.31	0.82	0.5
		β			0.56	0.39	0.38	0.1

		β			0.50	0.39	0.38	0.2
Berl Saddles	Ceramic	α			1.16		0.56	0.5
		β			0.47		0.25	0.1
Pall Rings	Plastic	α						0.2
		β						0.1
Pall Rings	Metal	α				0.43		0.1
		β				0.17		0.1
Raschig Rings in. Wall	Metal	α				1.20		
		β				0.28		
Raschig Rings in. Wall	Metal	α			1.59	1.01	0.80	0.5
		β			0.29	0.39	0.30	0.1
Equation								
Δp = Pressure Drop—in. H_2O/ft of packing								
G = Gas Mass Velocity—lb/sec ft^2								
L = Liquid Mass Velocity—lb/sec ft^2								
ρ_G = Gas Density—lb.ft^3								
α and β = Constants.								

ΔP is important to most column designs. Recognize that pressure drop will increase due to:

1. **Unsteady column operations**

2. Increased liquid/vapor loads
3. Breakage of ceramic packings (this can be serious)
4. Compaction/deflection of plastic packings.

14.11. Materials of Construction

Careful consideration should be given to fluids, temperatures of systems, aeration etc., likely to be encountered in service. Plastic may be attractive for the application but its *long-term heat deflection* characteristics needed to be determined. With time, many plastics will deform, thereby changing the packing bed characteristics, and hence the column pressure drop. As a general rule, do not select a plastic to operate within 50°F of its softening or deflection temperature.

14.12. Particle versus Compact Preformed Structured Packings

Particle packings (random) are usually (although not always) less efficient than the pre-packaged/preformed assemblies. However, particle packings are generally more flexible in loading and have better ability to handle “dirty” fluids.

The cost of the packing and its effect on system costs must be considered, as some packings are much more expensive than others, but produce very little improvement. Table 14-17 presents some comparisons information. The most common packings, and hence the ones with the most available data, are Raschig rings, Berl saddles and Pall Rings® (Norton Co.).

As a rough guide, the ratio of maximum random packing size to tower diameter is:

Raschig rings	1:20 (Reference 5)
	1:8 (Reference 52)

Berl saddles	1:10
lntalox saddles:	1:8 to 1:10 (Reference 6)

The 1:8 ratio is in more common use for most packings, however recent data indicated that Raschig rings require a larger ratio; approaching 1:20.

These ratios are useful in dealing with small towers, and serve as guides for somewhat larger ones. There are no guides detailing the smallest sized packing to place in a tower, but ½-in. is about the smallest ceramic in use with ¾-in. and l-in. being the most popular. Operating and pressure drop factors will usually control this selection.

Packed towers are not limited to small units; in fact the largest processing towers for absorption and stripping operations are probably of this type. Some units are 40 and 50 ft in diameter, and use 2-in. and larger packings to packing bed heights of 20–30 ft. Other units are 5–6 ft in diameter with 60 ft of packing. Towers of 24-in. dia. or smaller are most commonly packed because of the mechanical limitations of trays in small towers.

14.13. Fouling of Packing

Random and structured packings are susceptible to surface fouling due to process conditions and/or the presence of oxygen. Some systems will precipitate solids due to temperature and concentration effects. Bravo [135] discusses air–water stripping, and illustrates the effect of iron that is present in these systems on packing and fouling of distributors.

Fouled packing can significantly reduce the performance of the system checked for and should be examined when packed tower systems deteriorate. For some systems, an acid or solvent flush (or

wash) will accomplish the necessary thorough cleaning. The need for such maintenance should be recognized in initial system design. Also for water-soluble systems, the use of pH adjustment, pretreating with sequestering agents or biocides, ozone treatment and other steps specific to the system's problems [135] may be necessary or helpful.

14.14. Minimum Liquid Wetting Rates

To feed enough liquid into the tower to place a wet film of liquid effectively over all the packing, a minimum wetting rate (MWR) has been evaluated for guidance in operation and design. Morris and Jackson [52] recommend the MWR shown in Table 14-25.

Table 14-25 Minimum Wetting Rate*.

Packing	
Rings (Raschig, Lessing, etc.), thru 3-in. dia.	0.85
Grid type (wooden, etc.) (pitch 2 in.)	0.85
All packings larger than 3 in.	1.3
Polished metal packings and poor wetting surfaces (some plastics, glazed porcelain, etc.)	1.3, estimate to 2.5 Preferably etch surfaces to reduce problem.
*Compiled by permission from Morris and Jackson, *Absorption Towers*, Butterworth Scientific Pub. (1953) London and Imperial Chemical Industries, Ltd., Ref. 52.	

A minimum liquid rate for any tower packing is used by some designers as 1,500 lb liquid/(hr) (ft^2 tower) with reference to the liquid of properties of water.

where

L_{min} = liquid rate, $ft^3/(hr)$ (ft^2 crosssection)

MWR = value of minimum wetting rate from Table 14-25

a_t = specific surface area of packing, ft^2/ft^3

The minimum wetting rate is a function of the packing material surface (Table 14-25a) and the physical properties of the liquid involved, particularly its viscosity and surface tension. Most plastic, and some metal packings require surface treatment before they will wet uniformly or at all. Without film forming characteristics on the surface, the contact between liquid and vapor will be poor and performance can be expected to decline. The packing should be tested for wettability in service before selection.

Table 14-25A Packing Wetting Rates Related to Packing Material Surface.

Surface	**MWR_T Gpm/ft^2**	**Reasonable minimum wetting rate**		**Materials**
		ft^3/ft^2 hr	**m^3/m^2h**	
Unglazed ceramic	0.187	1.5	0.5	Chemical stoneware
Oxidized metal	0.27	2.2	0.7	Carbon steel, copper
Surface treated metal	0.40	3.2	1.0	Etched stainless steel
Glazed ceramic	0.80	6.4	2.0	
Glass	1.00	8.0	2.5	
Bright metal	1.20	9.6	3.0	Stainless steel, tantalum
PVC, CPVC	1.45	11.6	3.5	**

PVC-CPVC	1.45	11.0	3.5	**
Polypropylene	1.60	12.8	4.0	**
PTFE/FEP	2.00	16.0	5.0	**
* Modified by E.E. Ludwig from Glitsch literature.				
**Requires proper surface wetting treatment. Important that surface wetting be tested and treatment applied if necessary. Values based on > 43 ft^2/ft^3 specific area of packing. By permission of Mass Transfer, Inc., Bull. TP/US/B1 (1978) and Glitsch Bull. No. 345.				

Kister [90] has evaluated Schmidt's [92] somewhat complicated equation for minimum wetting rate and proposes:

(14-16)

where

MWR_G = minimum wetting rate, gpm/ft^2, generalized for other packings using Kister's evaluation

MWR_T = minimum wetting rate, gpm/ft^2 from Table 14-25A

a_t = specific surface area of packing, ft^2/ft^3

Another expression of “reasonable minimum wetting rate” [48] is given in Table 14-25A. The surface characteristics of the packing material are important in the type of liquid film (or droplets) that flow across, around, and drip off of its surface. The better the specific liquid wets the packing surface and forms a moving film, the more efficient will be the packing for distillation, absorption, *etc.* In general, it can be noted from the table that the surfaces that tend to wet easily have the lower minimum wetting rates. The data given in Table 14-25A do not agree well with the recommendations of Table 14-25, however. Although there is no validation, it is believed that the information in Table 14-25A is more current and represents a more

recent evaluation of available data. However, the fact that the results are not identified by packing design types suggests further evaluation of this factor is required.

Note that when packing is changed from one material to another, a new minimum wetting rate will be applied to the system.

14.15. Loading Point–Loading Region

Examination of Figure 14-20 shows the pressure drop of the packed bed with gas flow and no liquid flow as the dry curve. As liquid is added to the top of the packing, the effect on pressure drop is immediately noticeable. Note that the lower part of all the liquid rate curves are parallel to the slope of the "dry" bed curve. However, at a point a noticeable change in the slope of the pressure drop curve occurs. This is attributed to liquid hold-up in the bed changing from being only a function of liquid flow to being dependent on both liquid and gas flows. Although the change seems to occur for some packings at a distinct point, it is difficult to determine this accurately for all materials, and is perhaps better to consider a region − from the first point of inflection of the curve to its second. Towers are usually designed to operate with gas–liquid rates in the loading region or within about 60–80% of its upper point. As will be discussed later, the deviation from the loading point is greater for some situations, due to the relative proximity of the loading to the flooding point.

Figure 14-20

Pressure drop flow characteristics in conventional packed towers.

Reproduced by permission of the American Institute of Chemical Engineers, Sarchet, B. R., *Trans. Amer. institute of Chemical Engineers,* V. 38, No. 2 (1942) p. 293; all rights reserved.

For Figure 14-21A the loading region is centered about the 0.75 in/ft pressure drop curve; the preferred design range being 0.35 to a maximum of 1.0 in. of water/ft.

Figure 14-21A Sherwood-type correlation for flooding gas rate at a given liquid rate.

Used by permission of Zenz, F. A., *Chemical Engineering*, Aug. (1953), p. 181; all rights reserved.

Figure 14-21D indicates the loading region is centered about line B, which is a reasonable upper design condition.

Figure 14-21B and Figure 14-21C are the earliest generalized pressure drop correlations (GPDC) proposed, and they have been used for many industrial plant design. Progressively, Figure 14-21EFigure 14-21FFigure 14-21G and Figure 14-21H are more recent correlations. These charts will be discussed in a later section.

Figure 14-21B

Recent generalized pressure drop correlation. Note: ν = viscosity of liquid in centipoise.

Used by permission of Norton Chemical Process Products Corp.

Figure 14-21C

Generalized pressure drop correlation essentially equivalent to Figure 14-21B.

Used by permission of Norton Chemical Process Products Corp.

Figure 14-21D

Loading, flooding and pressure drop correlation (one of earlier versions).

Adapted by permission from Leva, M. *Tower Packing and Packed Tower Design*, 2nd ed. U.S. Stoneware Co.

Figure 14-21E

Latest version, generalized pressure drop correlation (GPDC).

Used by permission of Norton Chemical Process Products Cow. (rev. 1985).

Figure 14-21F

Strigle's latest generalized pressure drop correlation. Note G∗ = gas mass velocity, Ib/ft^2-sec.

Used by permission of Strigle, R. F. Jr., *Packed Tower Design and Applications; Random and Structured Packings*, 2nd ed. Gulf Publishing Co., © (1994) p. 19.

Figure 14-21G

Generalized pressure drop correlation for nonfoaming systems for IMTP metal random packing. Parameter of curves is pressure drop in inches of water/foot packed height. Numbers in parentheses are mm of water/meter of packed height.

Used by permission of Norton Chemical Process Products Corp., Bull-IHP-1, 12/91 (1987).

Figure 14-21H

Updated generalized pressure-drop correlation rearranged version of earlier Eckert and Leva, using linear scale for the ordinate and use of capacity factor, C_s.

Used by permission of Strigle, R. F., Jr., *Packed Tower Design and Applications; Random and Structured Packings*, 2nd ed. © Gulf Publishing Co. p. 21 (1994). Note: G = gas, Ib/ft^2-hr, L = liquid, Ib/ft^2-hr.

Figure 14-21I Norton IMTP Packing, Efficient Capacity Correlation for random metal packing only for nonfoaming systems. Norton recommends designing up to 90% of efficient capacity:

Used by permission of Norton Chemical Process Products Corp., Bull. IHP-1 (9/87).

Figure 14-21F is the most current version of the GPDC as presented by Strigle [139] to facilitate interpolation of the ordinate and pressure drop curves on the chart. The flooding and loading regions are not identified. For this chart:

(14-17)

(14-18)

(14-18A)

where

C_s = capacity factor, ft/sec

V_g = superficial gas velocity, corrected for densities ft/sec

F = packing factor from Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G

Updated by permission from R.F. Strigle, Jr., *Random Packings and Packed*

Table 14-26A Packing Factors,* ft^{-1} Wet and

Snowflake™ data by permissi

Packing type	**Material**	**¼**	**⅜**	**½**
Intalox® IMTP®	Metal			
Hy-Pak™	Metal			
Super Intalox®				
Saddles	Ceramic			
Super Intalox				
Saddles	Plastic			
Pall Rings	Plastic			
Pall Rings	Metal		76*–81	
Intalox® Saddles	Ceramic	725	330	20(
Raschig Rings	Ceramic	1600	1000	58(
Raschig Rings	″	700	390	30(
	Metal			
Raschig Rings	″			
	Metal			41(
Bed Saddles	Ceramic	259		24(
Snowflake™	Plastic			

Packing factors determined with an air-water system in 30″ I.D. Tower.
Norton Chemical Process Products Corp., Bull. ISPP-IR, E/90; ∗ by permi
*By permission Norton Co., from data compiled in Norton Co. Laboratorie

Table 14-26B Koch Packing Factors*.

		Nominal packing size, in										
Packing type	**Material**	**¼**	**⅜**	**½**	**⅝**	**¾**	**1 or #1**	**1¼**	**1½**	**2 or #2**	**3**	**3½ or #3**
Flexisaddles	Plastic						30			20		15
Flexirings	Plastic†				78		45		28	22		18
Flexisaddles	Ceramic	600		200		145	98		52	40	22	
*By permission, Koch Engineering Co. Inc.												
†Use for plastic or metal.												

Table 14-26C Glitsch Packing Factors*.

		Nominal packing size, in.										
Packing type	**Material**	**¼**	**⅜**	**½**	**⅝**	**¾**	**1 or #1**	**1¼**	**1½**	**2 or #2**	**3**	**3½ or #3**
Ballast™ Ring	Metal						48		28	20		15
Ballast™ Ring	Plastic				97		52		32	25		16
Ballast™ Saddle	Plastic						30			20		15
*By permission, Glitsch Inc.												

Table 14-26D Glitsch Packing Factors for

Cascade® Mini-Rings

Used by permission: Glitsch, Inc., Bull. 345.

Size	Plastic	Ceramic	Metal
0	—	—	55
0A	60	—	—
1	29	—	40
1A	30	—	—
1.5	—	—	29
2C	15	38	22
2A	30	—	—
2B	18	—	—
2C	19	—	—
2.5	—	—	19
3	11	24	14
3A	12	—	—
4	—	—	10
5	—	18	8
7	—	15	—

Table 14-26F Nutter Ring™ Random Packing

Used by permission of Nutter Engineering Co., a Harsco Corp.; Bull. NR-2

Size no.	Pieces/ft^3	Ft^2/ft^3	Lb/ft^3	% void	Packing factor**	Relative HETP values
0.7	4,740	69	11.0	97.8	xx	0.72
1.0	1,900	51	11.1	97.8	30	0.83
1.5	760	38	11.3	97.8	24	0.94
2.0	385	29	10.8	97.9	18	1.00
2.5	250	25	9.0	98.2	xx	1.18
3.0	120	20	8.3	98.4	xx	1.40

*Nutter use their own proprietary computer program and not the conventional GPDC Chart shown in Figures 9-21B–E.

**Values shown developed by Kister [90] from data supplied by Nutter Engineering Co., a Harsco Corp.

Table 14-26F Packing Factors (PF) (Dumped Packing)

(Source: Gas Processors Suppliers Association (GPSA), 12th Edition 2004, GPSA, Tulsa, Oklahoma, Used by permission of GPSA, All rights reserved).

		Nominal Packing Size (inches)										
Packing Type	**Material**	**1/4**	**3/8**	**1/2**	**5/8**	**3/4**	**1**	**1.25**	**1.5**	**2**	**3**	**3.5**
IMTP®	Metal				51		40		24	18	12	
Hy-Pack ™	Metal						45		29	26		16
Super-Intalox Saddles®	Ceramic						60			30		
Super-Intalox Saddles®	Plastic						40			28		18
Pall Rings	Plastic				75		55		40	26		17
Pall Rings	Metal				70		56		40	27		18
Intalox Saddles®	Ceramic	725	330	200		145	92		52	40	22	
Raschig Rings	Ceramic	1600	1000	580	380	255	155	125	95	65	37	
Raschig Rings	1/32” Metal	700	390	300	170	155	115		95			
Raschig Rings	1/16” Metal			410	300	220	144	110		62	32	
Berl Saddles	Ceramic	900		240		170	110		93	45		
Flexiring	Metal						49		65	23		16
Fleximax	Metal						35		26	17	11	
Cascade Mini Rings	Metal						39	36	33	26	18	
Cascade Mini Rings	Plastic						44		33	20	17	

Note: Values in this table are average values for the packing factor (PF). PF is actually a function of loading. Specific correlations for each packing

from the vendors

Table 14-26G Packing Factor Values (1/ft.) for Structured Packings (Source: Wankat, P.C., Separation Process Engineering 2nd Ed., Prentice Hall, 2007).

	Flexipac		**Gempak**		**Sulzer**		**Munters**		**Intalox**	
	2	4	2A	4A	CY	BX	1260	19060	2T	3T
F	22.6	6	16	32	70	21	27	15	17	13

L_h = liquid mass velocity, lb/(ft^2) (hr)
G_h = gas mass velocity, lb/(ft^2) (hr)
ρ_G = gas mass density, lb/ft^3
ρ_L = liquid density, lb/ft^3
ν = kinematic liquid viscosity, centistokes = centipoise/(ρ_L/62.4)

Note differences in some symbol units for various GPDC charts.

For tower sizing:

1. Calculate using ordinate value =
2. Calculate allowable gas mass velocity (ft/sec), C_S

Flow parameter at maximum flow location:

(14-19)

Strigle [139] reports that the correlation of Figure 14-21F (and probably Figure 14-21G, because it is from the same data) predicts

pressure drops to an accuracy of ±11%, and suggests that this is probably the best attainable with available data. Better accuracy can be obtained only when using data specific to a particular packing family of sizes, such as the Nutter, Norton, and Glitsch packings noted hereafter. For Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21EFigure 14-21FFigure 14-21G and Figure 14-21H, at high liquid rates, pressure drop may be somewhat greater than obtained from the charts (GPDC), particularly for smaller packing sizes. For higher liquid rates, the charts of the manufacturer of the packing should be used, rather than the GPDC. Strigle [139] reports that Kister and Gill's [93] tests indicate that from over 3,000 pressure drop measurements the results fit Figure 14-21G for 80% (excellent) and another 15% (reasonable) fit.

Strigle [82, 94] describes the hydraulics and HETP performance of a packed column by referring to Figure 14-22. As noted, the HETP values are essentially constant over a wide range of C_s values–shown as B-C on the figure. Note that C_s can be expressed as:

(14-20)

(14-20A)

Figure 14-22

Typical HETP curve illustrating operating and design relationships. Nomally reasonably constant over wide range of vapor flows. Note: C_S = capacity factor;

V = gas mass velocity = Ib/ft^2/sec.

Adapted by permission from Strigle, R. E Jr. and Rukovena, F., and reproduced with permission of the American Institute of Chemical Engineers, *Chemical Engineering Progress*, Mar. (1979); all rights reserved.

As vapor flow rate increases, so too does, the contact between liquid and vapor. Hence, the rate of mass transfer will rise, and, the HETP value will improve. With increasing vapor rate, liquid entrainment into the vapor phase will occur and lower the efficiency (and raise the HETP) to the "maximum operating efficiency" [94] at point F where the C_s value rises above the efficiency used for design. Thus, the "maximum operating capacity" is well below any physical flooding point. In fact, the term "maximum operating capacity" is considered as a much more meaningful term in establishing performance than "loading point" where earlier this was referred to as about point C [82]. The value of C_s at point D for atmospheric distillations has been found to occur at about 91% of the "maximum operating capacity" [94] at point F. The capacity factor C_s for design at point E has been set at the "maximum operating capacity," point F. The value of C_s for point E is approximately 87% of C_s at the maximum efficiency, point D. By setting the design capacity, C_s, as previously noted, the system should then be capable of operating up to 125% of design capacity and remain stable, and be conservative for mass transfer efficiency for vapor boil-up rates from point E to point F.

Thus, for mass transfer performance design a specific design HETP value should be established, which in effect represents the range from point B through E for C_s values above point F, the HETP values will be greater (and thus less efficient contact).

Some recent evaluations of data by other investigators indicate that a

so-called “loading region” does not exist as clearly as may be suggested by other studies, and therefore they suggest operating essentially up to the flooding point. For a good, reliable design that must allow for fluctuations in feed, and possibly column back-pressure upsets, designing to the flood region cannot be recommended. Design limits are discussed later.

14.16. Flooding Point

At the second sharp change in the slope of the pressure drop curve, Figure 14-20, the packing tends to hold up more and more liquid as the gas flow increases. This creates a rapid increase in pressure drop. The flooding point of the system is said to be the point of the second inflection of the pressure drop curve. Here, the liquid buildup on top of the packing increases and the pressure drop essentially becomes infinite for a finite increase in gas rate. In many actual cases, operation can be maintained at the flooding point, but it will be erratic, the performance (efficiency) of operation poor, and the entrainment carryover excessively high. It is obvious that towers are not designed for flood point operations, but at 40−60% of gas and liquid rates associated with this point. Figure 14-21D indicates that the flooding region is usually above 2.5 in. water/ft, but note how cramped and much more sensitive this condition becomes as the extreme right hand side of the graph is approached.

Kister's [93] study indicates that flooding with the newer packings occurs just below 2 in. water/ft of packing, somewhat below the prediction given by the earlier Eckert chart, Figure 14-21C and Figure 14-21D. The data plotted by Kister indicate that for larger packings (2 in. and 3 in., for example) the flooding point is found to be at much lower pressure drops than suggested by Figure 14-21C. The pressure drop at flood point has been found to be independent of the Flow Parameter on the charts, but does vary with the packing family and packing types [93, 95].

Kister [93] has developed a new approach to establishing the flood point that appears to fit the available data and is significantly more accurate than reading the upper curve on Figure 9-21C.

(14-21)

where

ΔP_{flood} = pressure drop at flood point for all random packings, in. water/ft of packing

F_p = packing factor, empirical, based on packing size and shape. 1/ft, see Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G.

By calculating the ΔP_{flood}, the capacity parameter can be determined using the calculated flow parameter and Figure 14-21E, and, if available, the SLE (Sherwood, Leva, Eckert) charts of Reference 93. For $F_p > 60$, the calculated ΔP_{flood} using the equation coincides with Eckert's (Figure 14-21C) flood line. Figure 14-23 illustrates the relationship between packing factors and flooding pressure drop, and is represented by the ΔP_{flood} equation. The predicted results of the equation within ±10 to 15% of the plotted data [93].

Figure 14-23

Flood pressure drop vs. packing factor for random packings.

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14.17. Foaming Liquid Systems

For an accurate design, the effects of foaming of the liquid as it flows through the random packed/structured bed should be known, estimated, or determined by experimentation. There is little published data on the subject except Eckert [24] and Strigle [82]. Generally, foaming systems produce higher pressure drops than nonfoaming, most probably due to the blocking of packing voids by the foam. Therefore, it is wise to determine the foaming behaviour of the specific system, and due to the lack of published data, generous allowance for pressure drop should be made in design, perhaps even 2 to 3 times greater than expected.

Hsu [124] presents equations for directly calculating random packings based on published data and which are adaptable for computer programming. These are very useful for studying the effects of variables. The basic data are essentially a match with Figure 14-21D.

14.18. Surface Tension Effects

Strigle [82] reports that there is no broadly documented agreement of the surface tension effects on the capacity of packed beds. Eckert [24, 82] concluded that surface tension of a nonfoaming liquid had no effect on capacity. These results were later confirmed. For absorption systems these results also hold [82].

For hydrocarbons in high-pressure fractionators Strigle [82] reports there is aeration of the rather low surface tension liquid phase. This effect increases as surface tension increases and as the vapor density increases. This reduces the effective liquid density, and so the volume occupied by a given mass of liquid is increased. This aeration effect can vary from 0.9 at atmospheric pressure for nonfoaming liquids to 0.7 for hydrocarbon systems (not absorbers operating at 35% of

critical pressure).

Kaiser [140] presents a correlation analysis for flooding in packed towers based on single phase hydraulics. However, Kaiser's article does not present a comparison of results with other more conventional techniques.

14.19. Packing Factors

The use of “packing factors” is well established in the design concepts of evaluating packed tower performance. Essentially all of the manufacturer's published data are for “wet and dumped” packing factors, F. Robbins proposes using only “dry” factors. This factor is a unique characteristic of each packing size and style/design. These factors cannot be determined by calculation from the physical dimensions; they are determined experimentally.

Packing factor selection significantly affects the performance of a packed tower system. These factors are only suitable for discreet particle type packing, and their values vary with the way the packing is installed in the tower. For example, the factors for a ceramic packing are different in the case of packing floated (dumped) into a tower full of water and the particles allowed to float down, compared to the same packing dumped into a dry empty tower (where significant breakage can occur and consolidate the packing), or even to packing “hand-placed” or stacked dry.

Often it is only necessary to change a packing size or type to modify the capacity and/or contacting efficiency of an existing tower, because this change affects the packing factor. Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G present specific packing factors from the manufacturers.

Many of the packings of the various manufacturers are essentially

identical in shape, size, and performance factors. Some packing manufacturers suggest adjusting packing factors for vacuum and pressure distillations; however, this should only be done after consultation.

The experimentally determined packing factors are the only reliable values to use for design calculations; although estimates can be made for packing shapes when no data are available. The packing characteristic is expressed as:

(14-22)

where

a_t = specific surface of packing, ft^2/ft^3

a = effective interfacial area for contacting, ft^2/ft^3

ε = fractional voids

The values of a/ε^3 determined experimentally by Lobo *et al.* are indicated [47]. These are the values in the development of the basic relation expressed in Figure 14-21A with correction of Ψ^2 suggested by Leva [41]. These a/ε^3 values were found to correlate with a considerable amount of the literature data, to within 12%. This would mean about a 6% error in tower diameter determined at flooding conditions.

Lobo *et al.* [47] proposed the packing factor, F, and experimentally determined that it better represented the data than the calculated a/ε^3 term. Values calculated using surface area per cubic foot and percent free gas space from manufacturer's tables can be in error by as much as 40%. The values are dependent upon the method of packing the tower, i.e., dry dumped, wet dumped, or wet dumped and shaken. The latter condition may approximate the situation after a tower has been

running for some time and the packing has settled.

Experience definitely indicates that the packing factor, F, increases with hours of operation for ceramic materials up to some limit. This is due to settling, breakage, plugging, *etc.* For design of commercial towers, values of F should be increased from 15 to 75% for ceramic materials, over the values read from Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G. The percent increase depends upon the tendency of the shape to disintegrate into smaller pieces during operations-flooding, gas surging, *etc.* In general, circular shapes exhibit the least tendency to break up. As a reasonable value where data are available, the average of the wet dumped, and wet-dumped-and-shaken values for tower voidages is recommended.

Leva [40] has converted the data of Lubin into correction factors to apply to a non-irrigated bed pressure drop, in order to derive a pressure drop for a liquid–gas system in the loading to flooding range. In general, this does not appear any more convenient to use than Figure 14-21D.

Relations expressing the fractional voids in a ring packed bed are useful in estimated the "ε" values for a/ε^3 determinations [47]. The average deviation is ± 2.6%.

Dry packed tower:

(14-23)

Wet packed, unshaken tower:

(14-24)

Wet packed and shaken tower:

(14-25)

where

l = ring height, in.
d_o = outside diameter of ring, in
d_i = inside diameter of ring, in.

The generalized correlations of Sakiadis and Johnson [59] are reported to satisfy a wide variety of systems.

Manufacturers of commercial packings provide packing factors for their products. Many (not necessarily all) of the commonly used packing factors are presented in Table 14-26A. These values are to be used with the application of Figure 14-21B and Figure 14-21F. Factors presented in Figure 14-24A and Figure 14-24B can also be used where the design requires them.

Figure 14-24A Packing factors (stacked packing selected grids).

Used by permission of Morris and Jackson, *Absorption Towers* Butterworth Scientific Publications, and Imperial Chemical Industries, Ltd., and adapted by U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

Figure 14-24B

Packing factors (screen packing and random dumped packing).

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

14.20. Recommended Design Capacity and Pressure Drop

The relationships in packed tower performance which are concerned specifically with the gas and liquid flows through a bed are expressed as a function of pressure drop. Pressure drop may be created by poor packing arrangements, i.e., tight and open sections in the bed, breakage of packing and settling of the bed, or plugging of void spaces by solids or reaction products. All of these occur in addition to the characteristic resistance of a particular packing to flow of fluids. This resistance will be different if the system is single phase rather than the two phase systems existing in most distillation, adsorption, scrubbing, or desorption operations. The basic pressure drop performance pattern of nearly all packings follows the same form as Figure 14-20.

Below the loading region, the pressure drop can be read from appropriate system curves if available, as Figure 14-20. However, for general use the data have been well correlated, Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21E and Figure 14-21F. The slope of most of the pressure drop curves indicates a proportionality of 1.8 to 2.8 power of the superficial gas mass velocity up the tower. This performance is typical of the gas-continuous range of packed tower

operation. In Figure 14-20, the curves for performance at water rates of L′ = 1,250 through L′ = 9,950 all seem to be a part of the same family. The curve L′ = 12,500 will be discussed later. The following discussion is valid only for gas-continuous operations. Fortunately the majority of packed towers operate in this condition. Liquid-continuous system will be considered later.

Pressure drop data for several styles and arrangements of drip point grid tile are given in Figure 14-25AFigure 14-25BFigure 14-25C and Figure 14-25D. These are not included in the general GPDC correlations for random packings.

Figure 14-25A Pressure drop in inches of water per foot of height, drip point tile, shape 6295 with crossed flue arrangement.

Used by permission of General Refractories Co.

Figure 14-25B

Pressure drop in inches of water per foot of height, drip point tile shape 6897 with continuous flue arrangement.

Used by permission of General Refractories Co.

Figure 14-25C

Pressure drop in inches of water per foot of height, drip point tile shape 6897 with crossed-flue arrangement.

Used by permission of General Refractories Co.

Figure 14-25D

Pressure drop in inches of water per foot of height; drip point tile shape 6146 with cross flue arrangement. Used by permission of General Refractories Co. **Figure 14-25E** Pressure drop in inches of water per foot of height, drip point tile shape 6146 with continuous flue arrangement.

Used by permission: General Refractories Co.

Figure 14-21EFigure 14-21F and Figure 14-21H show the latest general purpose correlations presented by several manufacturers of packing materials. The relative differences between the various correlations appear to be minor, thereby allowing packing performance to be evaluated on any chart, as long as the packing factors, F, have been determined on the same basis. Packing factors are presented in Table 14-26A and Table 14-26B and relate to discrete particle packings similar to those illustrated in Figure 14-6AFigure 14-6BFigure 14-6CFigure 14-6DFigure 14-6EFigure 14-6FFigure 14-6GFigure 14-6HFigure 14-6IFigure 14-6JFigure 14-6KFigure 14-6LFigure 14-6MFigure 14-6N(a)Figure 14-6N(b)Figure 14-6OFigure 14-6PFigure 14-6QFigure 14-6RFigure 14-6SFigure 14-6TFigure 14-6UFigure 14-6VFigure 14-6W and Figure 14-6X. The compacted,

structured, or grid packing materials such as shown in Figure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EEFigure 14-6FFFigure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MMFigure 14-6NNFigure 14-6OOFigure 14-6PPFigure 14-6QQFigure 14-6RRFigure 14-6SSFigure 14-6TT and Figure 14-6UU do not use the same packing factor concept for design evaluation. Therefore, the respective manufacturer should be consulted when using proprietary materials such as those in Figure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EEFigure 14-6FFFigure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MMFigure 14-6NNFigure 14-6OOFigure 14-6PPFigure 14-6QQFigure 14-6RRFigure 14-6SSFigure 14-6TT and Figure 14-6UU since they have their own design/rating technique.

14.21. Pressure Drop Design Criteria and Guide: Random Packings Only

A. Determine the height of packing required (not a function of diameter) from HETP in distillation section and Kga in absorption section. This provides the total height of packing required, considering the packing efficiencies and minimum wetting requirements previously discussed. From this total height, the expected total column pressure drop can be established, recognizing the pressure loss through support plates, distribution devices, *etc.*

B. Calculate the abscissa of Figure 14-21F; for example, or other abscissa values.

C. Select a design/operating pressure drop, as shown on the curves of Figure 14-21F. Suggested selection basis is as follows:

1. Low to medium pressure column operation; select design pressure drop of 0.40 to 0.60 in. water/ft of packing height,

although some towers will operate at 1.5 to 2.0 in. water/ft. Select a design C_s above point C and below point D, in Figure 14-22.

2. Absorption and similar systems; select a pressure drop of 0.1 to 0.4 in. water/ft of packing, or 0.25 to 0.4 for nonfoaming systems. For "median" foaming systems, use a maximum of 0.25 in. water/ft for the highest loading rate [82]; see Table 14-27. The values in the table should be reduced for high viscosity fluids. At rates above 20, limit the gas rate to 85% of the rate which would give a pressure drop of 1.5 in. water/ft referring to the GPDC, Figure 14-21F.

Table 14-27 Maximum Recommended Liquid Loading.

Used by permission of Strigle, R. F., Jr., "Random Packing and Packed Towers" © (1987), Gulf Publishing Co., all rights reserved.

Packing size (in.)	**Liquid rate (gpm/ft^2)**
¾	25
1	40
1½	55
2	70
3½	125

3. Atmospheric or pressure distillation; select a pressure drop of 0.50–1.0 in. water/ft.

4. Vacuum distillation varies with the system and particularly with the absolute pressure required at the bottom of the column; for average systems select a low pressure drop; in the range of 0.1 to 0.2 in. water/ft of packing. For vacuum service–75 mm Hg and lower, the pressure drop is obtained from the GPDC, Figure 14-21F

and Figure 14-21G (for Norton's IMTP packing only). It can be up to 20% higher than industrial experience for the same C_s values at a flow parameter (FP) of 0.01 [82]. At an absolute column pressure of 10 mm Hg or less, the actual pressure drop can be up to 30% lower than that read from the Figure 14-21F, at the FP of 0.01 and the same C_s value [82]. The conclusion is that the generalized correlation –Figure 14-21F and Figure 14-21G– always give a conservative design ΔP at operating pressures of less than 70 mm Hg abs. Several other factors must be considered, such as variability of gas and liquid rates and densities or specific volumes.

5. Foaming materials should be operated at 0.1−0.25 in. water/ft, or better, obtain data. Actual ΔP/ft may run 2-5 times chart values.

6. Vacuum service requires recognition of minimum liquid flow; refer to section on minimum wetting rate. Pressure drops are designed to be low, but normally not lower than 0.10 in. water/ft.

7. Stripping gas from a liquid phase; pressure drop is usually set to be 0.15 to 50 in. water/ft packing at the flow rates for maximum loading, or maximum operating capacity. For foaming systems set the pressure drop at ≤0.25 in. water/ft using inert gas. For steam use, set the pressure drop ≤0.3 maximum in. water/ft for medium foaming conditions [82].

8. For high pressure distillation of light hydrocarbons, industrial performance indicates that the pressure drop actually obtained is about twice that predicted by the use of the GPDC charts, Figure 14-21F and Figure 14-21G (for Norton's IMTP packing only). When the vapor density is at least 6% of the liquid density, the actual pressure drop is expressed [82]:

(14-26)

where

C_s = capacity factor, ft/sec

σ = surface tension, dynes/cm

F = packing factor

For non-hydrocarbon systems Strigle [82] recommends the general outline to follow, but more detail for a specific design maybe obtained from the reference as well as the manufacturer:

1. Select a design C_s vapor rate as a percent of the MOC (maximum operating capacity) of the packing.
2. Adjust maximum C_s required, considering:
   a. Effect of liquid viscosity: The maximum operating capacity varies as the 0.2 to 0.25 power of surface tension of the liquid phase.
   b. Packing capacity increases as the reciprocal of liquid viscosity to the 0.1 to 0.13 power. This effect is limited to liquid viscosities not <0.09 cPs.
   c. Select design C_s at 80 to 87% of the MOC, C_s for the packing.
   d. When the gas density exceeds 6% of the liquid density, the pressure drop should be examined by Equation 14-26.
   Pressure drop should not exceed (at maximum operating capacity) [94] [139]:

(14-27)

or, from Kister and Gill [93, 139], pressure drop at flood for

random dumped packing:

(14-21)

e. For pressures approaching critical value, contact the packing manufacturer for performance guidance.

f. Select a design C_s that provides allowances for operational and capacity fluctuation and/or surges. 15–25% above the design rate is usually adequate [82], making the design C_s 80–87% of the maximum operational C_s for the specific packing. The maximum operational capacity, MOC, $C_{s,max.}$ is the point, C (Figure 14-22), where the vapor rate has increased to a point where it begins to interact with the liquid. In previous publications, this point was termed “loading point". Actually for rates slightly greater than point C, the efficiency of the packing increases (HETP becomes smaller).

3. Select a packing and determine its packing factor from Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G. Packing is selected for its expected process HETP or Kga performance, pressure drop and materials of construction for the system. Table 14-17 presents summary comments for applications. Selection guides are as follows:

a. As packing factor, F, becomes larger by selection of smaller sized packing; the gas capacity of the column is reduced; and pressure drop will increase for a fixed gas flow.

b. Some packings are sized by general dimensions in inches, while some shapes are identified by numbers, #1, #2, #3 for increasing size.

c. Not all packings are manufactured in all materials of

construction.

d. Packing size versus tower diameter recommendations; general guides not mandatory, base selection on performance.

Tower Diameter, ft	Nominal Packing Size, in.
<1.0	<1
1.0–3.0	1–1½
>3.0	2–3

Table 14-27 shows what Strigle [82] recommends for the maximum liquid loading as related to packing size.

4. Referring to Figure 14-21B and Figure 14-21C, read up from the abscissa to the pressure drop line selected, and read across to the ordinate (note differences):

(14-28)

(14-29)

(14-30)

Note units change for Figure 14-21G, where:

(14-31)

υ = kinematic liquid viscosity, centistokes

Substitute F and the other known variables into the equation and solve for G, the gas mass flow rate, lbs/ft^2 sec, or G_g, lb/hr-ft^2, as applicable.

Then, determine the required tower crosssection area and diameter:

14.22. Effects of Physical Properties

For nonfoaming liquids, the capacity of a packing is independent of surface tension. Foaming conditions reduce capacity significantly, and design should recognize this by selecting an operating pressure drop of only 50% of that for a normal nonfoaming liquid.

For liquids of viscosity of 30 centipoise or lower, the effect on capacity is small. For high viscosity liquids select larger packing to reduce pressure drop, and also consult the packing manufacturer.

Robbins' [96] correlation for pressure drop in random particle packed towers is based on a “dry packing factor,” F_{pd}, whereas most manufacturer's published values derive F (packing factor) from wet and dumped data, such as is used in Figure 14-21AFigure 14-21BFigure 14-21CFigure 14-21DFigure 14-21EFigure 14-21F and Figure 14-21H. Reference to the tables in Robbins' publication indicates that the differences between the dry, F_{pd}, and the wet and dumped, F (such as Table 14-26A) may be small, being from 0 to 10 points, averaging about 2-3 points lower. The packing manufacturer should be consulted over what dry packing factors to use in Robbins' method. The “dry” data simply means that there is no liquid (but gas)

flowing. Robbins [96] lists values of F_{pd} for metal, plastic and ceramic packings.

Dry bed pressure-drop [96]:

(14-32A)

Values of C_o come from Leva [41].

Robbins' new equation for generalized pressure drop for random tower packings:

(14-32B)

(14-32C)

(14-32D)

The method as described by Robbins [96] is as follows:

1. For operating pressures above 1 atm, multiply G_f (Equation 14-32C) by .
2. For small packings with F_{pd} of 200 or greater, substitute $\mu^{0.2}$ in previous equation for $\mu^{0.1}$
3. For large packings with F_{pd} below 15, use $(20/F_{pd})^{0.5}$ in place of $[F_{pd}/20]^{0.5}$ in previous equation for L_f, Equation 14-32D.
4. Dry bed pressure drop:

(14-32E)

Dry bed pressure drop values are usually between 0.1 and 0.5 in. water/ft of packing [96]. Use Equation 14-32B when L_f is below 20,000. Packings operate essentially dry when L_f is below 1,500 (about 3 gpm/ft^2) at F_p = 20. Pressure drop at flooding is suggested to be predicted by Kister and Gill's relationship [93] presented in this text.

Robbins [96] suggested random packed column design is similar to others presented in this text, but highlighted to determine diameter of packed column:

1. Establish liquid and vapor rates.
2. Determine fluids physical properties.
3. Select design pressure drop for operations. Suggested values of below 1.0 in. water/ft. Low-pressure, atmospheric, and pressure columns usually require 0.5 to 0.7 in. water/ft, with absorbers and strippers around 0.2–0.6 in. water/ft. For vacuum distillation low values of 0.05–0.6 in. water/ft are often necessary, usually depending on the required boiling point of the bottoms.
4. Calculate L_f/G_f, values of F_{pd} cancel out.
5. Using Equation 14-32B and trial and error, calculate using G_f, for given L_f/G_f until the desired ΔP is obtained. L_f must be below 20,000. For higher L_f use chart in the original article (not included here).
6. Select packing for column, and establish packing factor, F_{pd}.
7. Calculate column crosssectional area using the operational gas rate, G, and the calculated value of G_f (gas loading factor);

(14-32F)

8. Establish tower diameter; Robbins [96] recommends that the tower diameter should be at least 8 times the packing size; if not, repeat the calculations with different packing.

where

A = tower crosssectional area, ft^2
G_o = constant specific to a particular packing
G_l = constant specific to a particular packing
$G_3 = 7.4 \times 10^{-8}$
$G_4 = 2.7 \times 10^{-5}$
D = tower diameter, ft
F_p = packing factor, dimensionless
F_{pd} = dry-bed packing factor, dimensionless
$F_s = V_s (\rho_g)^{0.5}$, (ft/sec) $(lb/ft^3)^{0.5}$
G = gas loading, $lb/hr\text{-}ft^2$
G_A = design vapor flow rate, Ib/hr
G_f = gas loading factor
L = liquid loading, $lb/hr\text{-}ft^2$
L_f = liquid loading factor
ΔP = specific pressure drop, in. water/ft of packing
ΔP_{pb} = specific pressure drop through dry bed, in. water/ft of packing
V_s = superficial gas velocity, ft/sec
ρ_g = gas density, lb/ft^3
ρ_L = liquid density, lb/ft^3
μ = liquid viscosity, centipoise

Strigle [82] and Kister [93] point out the importance of evaluating data (where available) to reduce the need for interpolating the GPDC

charts. The question of reasonably accurate (±10 to 15%) flooding pressure drop data has been studied by Kister [93], and the results suggest that the establishment of flooding pressure drop curves such as in Figure 14-21C may not be consistently accurate due to the variations in data used for correlation, i.e., data just as flooding begins, and then at full flooding. The data presented are for gas–liquid systems and not liquid–liquid extraction as illustrated by Strigle [82].

Strigle [82] identifies a regime 20% above point F on Figure 14-22 as the maximum hydraulic capacity, and is termed the flooding point for atmospheric operations.

Kister [93] has correlated large quantities of available data for flooding, and offers a new correlation based on the same GPDC correlation, *i.e.* the Sherwood-Leva-Eckert (SLE) correlation chart as developed by Strigle [82], Figure 14-21H, for semi-log plot. Kister [90, 93] has presented plots showing the data on the Strigle or SLE charts for a wide selection of packing. In effect this illustrates how the data fit the generalized charts.

As an example, Figure 14-26 and Figure 14-27 from Kister [93] illustrate the collected data superimposed on the SLE chart for the specified pressure drops, specific packing and column size, and packing heights at designated packing factors. Kister [93] recommends using the specific SLE chart (also see Kister [90] for a wide selection of charts) and interpolating and extrapolating the curves when the design/operating requirements fall close to the data points on the selected chart. Extrapolating too far can ruin the validity of the pressure drop results.

Figure 14-26

SLE Data Chart for 1-in. metal Pall rings, aqueous systems, pressure drop only. Data from 15–84 in. dia. test columns with packed heights of 2–10 ft.

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Figure 14-27

SLE Data Chart for 2-in. metal Pall rings, aqueous systems, pressure drop only. Data from 30–84 in. dia. test columns with packed heights of 2–18 ft.

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When the design/operating requirements are far from the chart's data points, contact the manufacturer of the packing for data, and also consider selecting another type and/or size of packing in order to provide a better fit. In selecting a chart to use, do not overlook the physical nature of the process system, i.e., whether predominantly aqueous or nonaqueous, and whether it involves foaming characteristics, or high or low viscosity fluids. Kister's assessment is that both the study in Reference 93 and Strigle [82] show that the large volume of published data does fit the Strigle [82] charts (Figure 14-21E and Figure 14-21F) quite well and gives “good pressure drop predictions.” A significant variation is that the curves predict an optimistic (too low) value for nonaqueous systems at high flow parameters of the chart. Also, similarly optimistic values are noted for

nonaqueous systems at low flow parameters, *e.g.* vacuum distillation. Unfortunately, most data

Table 14-28. Maximum Capacities of Various Packings.

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Packings	**C_s at maximum efficiency, ft/sec**	**ΔP at maximum efficiency, H_2O/ft**	**C_s at maximum capacity, ft/sec**	**ΔP at maximum capacity, H_2O/ft**
2 in. Pall Rings	0.295	0.81	0.315	1.29
2 in. Intalox Saddles	0.248	0.96	0.279	1.74
#50 Intalox Metal Packing	0.327	0.52	0.345	0.88
#40 Intalox Metal Packing	0.290	0.60	0.310	0.99
1-½-in. Pall Rings	0.269	0.95	0.287	1.48
1-½ in. Intalox Saddles	0.211	1.09	0.237	1.88
#40 Intalox Metal Packing	0.290	0.60	0.310	0.99
#25 Intalox Metal Packing	0.260	0.96	0.278	1.62
Note: #25 Intalox®, Norton = app. 1-in. size.				
#40 Intalox®, Norton = app. 1½-in size.				
#50 Intalox®, Norton = app. 2-in. size.				

were obtained on a small scale as compared with industrial equipment, and so the statistical fit is still not adequate to use with confidence in industrial design[93].

14.23. Performance Comparisons

Strigle [94] presents some helpful comparisons with reference to Figure 14-22 and Table 14-29Table 14-30 and Table 14-31 (from [94]) present key design and operational points related to Figure 14-22.

Table 14-29 Design Efficiency and Capacity for Selected Packings.

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Packing	Design C_s, ft/sec	Design HETP, ft
2 in. Pall Rings	0.256	2.32
2 in. Intalox Saddles	0.216	2.50
#50 Intalox Metal Packing	0.284	2.12
#40 Intalox Metal Packing	0.252	1.74
1-½-in. Pall Rings	0.234	1.78
1-½ in. Intalox Saddles	0.183	1.87
#40 Intalox Metal Packing	0.252	1.74
#25 Intalox Metal Packing	0.226	1.38
Note: #25 Intalox®, Norton = app. 1-in. size.		
#40 Intalox®, Norton = app. 1½-in size.		
#50 Intalox®, Norton = app. 2-in. size.		
See Figure 9-22 for C_s vs. tower internals.		

Table 14-30 Comparison of Maximum Capacity Designs.

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Packing	Relative tower diameter	Relative packed height	Relative packing volume
2 in. Pall Rings	1.00	1.00	1.00
2 in. Intalox Saddles	1.09	1.08	1.28
#50 Intalox Metal Packing	0.95	0.91	0.82
#40 Intalox Metal Packing	1.01	0.75	0.76

Packing			
1-½-in. Pall Rings	1.00	1.00	1.00
1-½ in. Intalox Saddles	1.13	1.05	1.34
#40 Intalox Metal Packing	0.96	0.98	0.91
#25 Intalox Metal Packing	1.02	0.78	0.80

Table 14-31 Comparison of Constant Pressure Drop Designs

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ΔP = 0.5 in water per therotical plate			
Packing	**Relative diameter**	**Relative height**	**Relative volume**
2 in. Pall Rings	1.00	1.00	1.00
2 in. Intalox Saddles	1.10	1.08	1.31
#50 Intalox Metal Packing	0.85	0.91	0.66
#40 Intalox Metal Packing	0.89	0.75	0.59
1-½ in. Pall Rings	1.00	1.00	1.00
1-½ in. Intalox Saddles	1.13	1.05	1.34
#40 Intalox Metal Packing	0.87	0.98	0.74
#25 Intalox Metal Packing	0.97	0.78	0.73

For design, select C_s above the loading point C (on Figure 14-22), and below the maximum efficiency rate, point D. Usually, the flow at point C is 70 to 75% of that at the maximum point F [94]. The design C_s (point E) should allow at least 15% increase in rates before

reaching the maximum operating capacity of the packing. This allows for normal variation in operations [94]. The design C_s value should be at least 80 to 85% of the maximum operational C_s (MOC) for the best turndown condition during operation.

14.23.1. Prediction of Maximum operating Capacity (MOC)

Strigle [94] proposed this term to better describe the performance of a packed column at or near the previously described "loading point". Kister [93] evaluated the limited published data, and proposed using the MOC at 95% of the flood point. The flood point can be estimated by Equation 14-21 or from the plots in References 90 and 93. The data are reported to be within 15–20% of the prediction [93]. See Figure 14-22 for the identification of MOC on the HETP vs. C_s chart. For more accurate information contact the respective packing manufacturers.

14.24. Capacity Basis for Design

Whether for a distillation, absorption, or stripping system the material balance should be established around the top, bottom, and feed sections of the column. Then, using these liquid and vapor rates at actual flowing conditions, determine the flooding and maximum operating points or conditions. Then, using Figure 14-21BFigure 14-21E and Figure 14-21F, establish the pressure drop, or assume a pressure drop and back-calculate a vapor flow rate, and from this a column diameter. Even though the column diameter may show a difference between the requirements at its top, bottom and/or middle, it is not necessary to design a column with varying diameter to deal with these theoretically varying requirements. Sometimes the calculations show that a larger diameter in the upper section for vacuum columns would give better performance. A constant diameter column should be thoroughly evaluated before pursuing this, since it is the least expensive option, and is somewhat easier to fit all the parts together. Even though the pressure drop or HETP may not be the

same throughout [154, 156, 157], adjustments can be made, if warranted, even by changing the packing size or style/type in various segments of the column. This requires some careful calculations for the effective HETP and the total height of the individual sections and then the total column/tower. But it is often worth the effort, particularly for tall towers, say above 20-30 ft of packing. For low pressures, as well as other columns, determine the pressure drop for each packed section of the column, plus the pressure drop through the internal components.

14.24.1. Flooding

In packed columns, flooding is caused when the energy potential of the vapor is greater than that of the liquid (i.e. when all the void spaces in the packing are filled with liquid), so that the liquid travels upwards and the column floods. This gives a ΔP of 2–3 in. H_2O/ft of packing depth, and a column cannot be operated successfully under such conditions. Flooding depends on a number of parameters, and empirical correlations of the variables have been presented by Bain and Hougen [161], Lobo *et al.* [47], Sherwood *et al.* [162] and Eckert [163].

Eckert's modification of Sherwood's flooding correlation has been widely applied to designing packed columns. Blackwell [164] developed mathematical models representing the generalized pressure drop curves excluding the flooding condition. Kessler and Wankat [165] formulated a model to represent Eckert's flooding curve within the range of the original data.

14.24.2. Operating and Design Conditions

Packed towers are typically operated at a gas velocity that corresponds to about 50–80% of flooding condition. This usually results in a ΔP of 0.5–1.0 in. H_2O/ft of packing. Alternatively, a large decrease in flow rate causes channeling through the packed bed, and could result in

poor column performance.

Two principal design methods have often been employed in sizing packed towers. One is the selection of an allowable ΔP, and the other is to base calculations on some fraction of the flooding capacity. Nguyen [166, 167], uses the latter, and employs Chen's least–squares method to determine the tower diameter at 50% of the superficial flooding rate. This means that the ΔPs are 2 in. H_2O/ft. of bed, or greater. Here, an allowable ΔP is selected to determine the diameter of a packed tower. Kessler and Wankat's model is employed to determine the vapor rate at 70% of the flooding condition.

The maximum recommended design ΔP is 1.0 in H_2O/ft of liquids with a specific gravity of 0.8 or more. A lower maximum design ΔP is recommended for less dense liquids. The recommended design ΔP criterion for sizing a packed tower should fall within the ranges shown in Table 14-32 [168]. Packed beds may be scaled up using the modified Sherwood correlation, as illustrated in Figure 14-21C. The packing factor for various packing types is as shown in Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G, because the capacity of the bed is independent of tower diameter and height. Packings that have sizes too large for the tower diameter will show a greater contact caused by the extra voidage, (i.e. where the pieces of packing contact the tower wall), and a lower ΔP than that predicted by their published packing factors. This effect reduces to zero for most packings when the ratio of packing size to diameter exceeds 10:1 [169]. This factor is necessary when data from a pilot plant tower, which is filled with large packings are used for full scale designs.

Table 14-33. Constants for Each Pressure Drop Correlation.

Source: Blackwell, W.W., Chemical Process Design On

Source: Blackwell, W.W., Chemical Process Design On A Programmable Calculator, McGraw-Hill Book Company, Inc., 1984.			
Pressure Drop in. H_2O/ft. of packed depth	**Constants**	**Pressure Drop in. H_2O/ft. of packed depth**	**Constants**
0.05	C_0 = -6.30253	0.50	C_0 = -4.39918
	C_1 = -0.60809		C_1 = -0.99404
	C_2 = -0.11932		C_2 = -0.16983
	C_3 = -0.00685		C_3 = -0.00873
	C_4 = -0.00032		C_4 = -0.00343
0.10	C_0 = -5.50093	1.0	C_0 = -4.09505
	C_1 = -0.78508		C_1 = -1.00120
	C_2 = -0.13496		C_2 = -0.15871
	C_3 = -0.00134		C_3 = -0.00797
	C_4 = -0.00174		C_4 = -0.00318
0.25	C_0 = -5.00319	1.5	C_0 = -4.02555
	C_1 = -0.95299		C_1 = -0.98945
	C_2 = -0.13930		C_2 = -0.08291
	C_3 = -0.01364		C_3 = -0.03337

	-0.01264		-0.03237
	C_4 = -0.00334		C_4 = -0.00532

Table 14-32 Generalized Pressure Drop Rules

Source: Branan, C., The Process Engineer's Pocket Handbook, Gulf Publishing Co., Houston, Texas, (1976).

Service	Design pressure drop in. in. H_2O/ft. of packed depth
Absorbers and Regenerators (Nonfoaming System)	0.25-0.40
Absorbers and Regenerators	0.10-0.25
Atmospheric or Pressure Stills and Fractionators	0.40-0.80
Vacuum Stills and Fractionators	0.10-0.40

Design Equations

The packed tower sizing method is based upon the generalized ΔP correlation as shown in Figure 14-21C.

The abscissa on the generalized ΔP curve is expressed as:

(14-33)

where and the ordinate is expressed as:

(14-34)

where

(14-35)

and $C_0 - C_4$ are constants derived for each ΔP. The values of these constants for each ΔP correlation are shown in Table 14-34.

(14-36)

The tower area is:

(14-37)

The tower diameter is defined by:

(14-38)

At flooding condition, Eckert's flooding curve is represented as:

(14-39)

where

(14-40)

Using Kessler and Wankat's model to represent the flooding condition,

(14-41)

and

Table 14-34. Input Data and Computer Results of Example 14-1.

DATA141.DAT			
3230.6	1136.0	0.0738	
62.3	155.0	1.0	1.0
PACKED TOWER DESIGN			
LIQUIED FLOW RATE, lb/hr.:	3230.6		
GAS FLOW RATE, lb/ft.^3:	1136.0		
GAS DENSITY, lb/ft.^3:	0.0737		
LIQUID DENSITY, lb/ft.^3:	62.3000		
TOWER PACKING FACTOR:	155.0		
PRESSURE DROP, INCH-H20:	1.000		
LIQUID VISCOSITY, cP:	1.000		
X COORDINATE IN THE GENERALIZED PLOT:	0.0979		
Y COORDINATE IN THE GENERALIZED PLOT:	0.0719		
CALCULATED VAPOR RATE, lb/s.td^2:	0.2617		
X COORDINATE AT FLOODING CONDITION:	0.0978		
Y COORDINATE AT FLOODING CONDITION:	0.1335		
VAPOR RATE AT 70% FLOODING CONDITION:	0.2496		
VAPOR VELOCITY AT 70% FLOODING CONDITION, Ib/s.ft^2:	3.39		
TOWER AREA, ft.:	1.21		
TOWER DIAMETER, inches:	14.87		
TOWER DIAMETER, ft.:	1.24		
TOWER AREA AT 70% FLOODING CONDITION, ft^2:	1.67		
TOWER DIAMETER AT 70% FLOODING CONDITION,			

inches:	17.49
TOWER DIAMETER AT 70% FLOODING CONDITION, ft.:	1.46
FLOOD PRESSURE DROP, in. H2O/ft. packing:	3.926

(14-42)

where

g_c = 32.174 $lb_m./lb_f$. ft/sec^2

μ_L = liquid viscosity, cP

At flooding condition,

(14-43)

The mass vapor rate (lb/ft^2.s) at 70% flooding condition can be expressed as:

(14-44)

The vapor velocity (ft/s) at 70% flooding condition is expressed as:

(14-45)

Since the flooding correlation is not perfect, a 95% confidence with a safety factor of 1.32 is used to determine the area at flooding condition (Bolles and Fair [172]).

Therefore, area of the tower at flooding condition is:

(14-46)

The tower diameter at 70% flooding condition is:

(14-47)

In the region below the flooding curves, the pressure drop can be correlated with an equation of the form:

The pressure drop at incipient flooding from Kister and Gill's [93, 139] equation is:

(14-21)

where

A = Tower area, ft^2, Equation 14-37

$Area_{flood}$ = Tower area at 70% flooding condition, ft^2, Equation 14-46

$C_0 – C_4$ = Pressure drop correlation constants.

D = Packed tower diameter, ft.

dia_{flood} = Tower diameter at 70% flooding condition, ft, Equation

14-47

F_{LG} = Flow parameter in Equation 14-40

F_p = Packing factor for packing material

G = Vapor rate, lb/hr.

G_1 = Vapor mass velocity, lb/(ft^2. sec. tower cross section), Equation 14-36

G_2= Vapor mass velocity, lb/(ft^2. sec. tower cross section), Equation 14-44

g_c= conversion factor, 32.174 $lb_m./lb_f$. ft/sec^2

L = Liquid rate, lb/hr.

L_1= Liquid mass velocity, lb/(ft^2 sec. tower crosssection)

X= Abscissa in the generalized pressure drop correlation chart.

Y= Ordinate in the generalized pressure drop correlation chart.

Y_1= Ordinate in Eckert's flooding curve

V_{GF}= Vapor velocity at 70% of flooding condition, ft/sec., Equation 14-45.

Greek letters

α, β= Constants in Tables 14-24 and 14-25

μ_L= Liquid viscosity, cP

ρ_L= Liquid density, lb/ft^3

ρ_G= Vapor density, lb/ft^3

Ψ= Ratio of water density to liquid density, ρ_{water}/ρ_L

υ= Kinematic liquid viscosity, centistokes, note:

ΔP= Pressure drop, in.H_20/ft. packing

Example 14-1

A packed tower with 1-inch Raschig rings is to be used to absorb ammonia from air, by contacting the gas stream with water. The

entering gas flow is 40 lb mol/h and contains 5.0% of NH_3. Ninety per cent of the NH_3 is to be removed from the air. The entering water flow rate is 3,200 lb/h. Absorption is to be carried out at 1 atm. Abs., and 20°C. Estimate:

1. The diameter of the tower for a ΔP of 1.0 in. H_2O/ft. packing.
2. The vapor rate at 70% of flooding condition.

Physical properties:

= 0.0737 lb/ft^3

Solution

When calculating the tower diameter, conditions at its base should be used, since this corresponds to the maximum flow rates.

At the bottom of the column:

Molecular weight of ammonia,

NH_3 removed from the air = (0.9) (0.05) (40)

Liquid flow rate at the bottom = 3200 + 30.6
= 3230.6 lb/h

Using the generalized pressure drop correlation of Strigle, Figure 14-21F, gives:

The flow parameter,

From Figure 14-21F at , the ordinate is 0.069

That is:

The liquid viscosity in centistokes,

= 1/(62.3/62.4) = 1.0016 cSt.

and the mass vapor rate, G_1 is expressed as:

Area of the packed tower:

Hence, tower diameter:

(b). Using Kessler and Wankat's correlation to represent the flooding condition [165],

and

At flooding condition:

The mass vapor rate (lb/ft^2. s) at 70% flooding condition can be expressed as:

The vapor velocity (ft/s) at 70% flooding condition is:

The tower area at 70% flooding condition, using a safety factor of 1.32 is:

Tower diameter is:

The pressure drop at incipient flooding from Kister and Gill's [93, 139] equation is:

(14-21)

Computer Program

A computer program PROG 141.for has been developed to calculate the diameter of packed towers for pressure drops of 0.05, 0.10, 0.25, 0.50, 1.0 and 1.5 in. H_2O/ft. The simulation is based on the generalized pressure drop correlation of Sherwood *et al.* [162]. The curves have been fitted to a series of equations developed by Blackwell [164]. In addition, the program calculates the vapor rate and vapor velocity at 70% of flooding condition, based on Kessler and Wankat's correlation, and uses a safety factor of 1.32 to determine the diameter of the column at 70% flooding condition. The pressure drop at flooding condition is determined using Kister and Gill's equation [93, 139]. Table 14-34 shows the input output data for the packed tower of Example 14-1.

14.25. Proprietary Random Packing Design Guides

Norton Intalox® Metal Tower Packing (IMTP®)

Norton offers a new, high performance system centered around an improved Intalox® metal tower packing, including effective internals which provide the distribution and pressure drop consistent with the higher performance of the packing itself.

Figure 14-6J and Figure 14-6U illustrate the IMTP packing. The manufacturer's key performance descriptions are [83]:

1. Greater capacity and efficiency than fractionation trays and other dumped packings.
2. Pressure drop approximately 40% lower than equivalent size Pall rings.
3. Low liquid hold-up.
4. Structural strength allows packing depths to 50 ft or more.
5. Easy to use in distillation, from deep vacuum where low pressure drop is beneficial, to high pressure where capacity surpasses many trays.
6. HETP values nearly independent of flow rate.
7. Packing properties related to a performance curve:
    a. The system base HETP of a packing, which is the flat HETP value produced by uniform distribution.
    b. The efficient capacity (not the same as hydraulic capacity or flood point) of a packing, which is the greatest vapor rate at which the packing still maintains the system base HETP. Norton [96] rates packings by percent of efficient capacity rather than percent of flood.

Capacity Correlation [83]

Figure 14-21G and Figure 14-21I present the proprietary estimating capacity charts for the various sizes of the Intalox® packing for a nonfoaming system. The system base HETP of a packing is the flat HETP value produced by uniform distribution, see Figure 14-22.

The terminology for the chart referenced to Norton's [83] Intalox® random packing of various sizes designated as:

Size No.	Dimensions, in.
15	-
25	1
40	-
50	2
70	-

Design information used by permission of Norton Chemical Process Products Co.

(14-48)

(14-49)

Capacity rating, C_o = feet/sec., or m/sec, from chart

(14-50)

This is the greatest vapor rate at which the packing still maintains the system base HETP [83].

Capacity rating = $[C_s/C_{sc}]$ (100) = % capacity [83].

Pressure drop equation: For IMTP packing, nonfoaming system, use

Figure 14-21G and Figure 14-21I.

where flow parameter, X = as previously noted, Equation 14-48.

Value for F coefficient (used by permission Norton [83])

IMTP size	No. 15	No. 25	No. 40	No. 50	No. 70
F when C_s in m/sec	549	441	258	194	129
F when C_s in ft/sec	51	41	24	18	12

where

ν = liquid kinematic viscosity, centistokes
= μ/(sp.gr.) centistokes
μ = liquid viscosity, centistokes
σ = surface tension, dynes/cm
L = liquid mass rate, lb/hr
G = gas mass rate, lb/hr
ρ_L = liquid density, lb/ft^3
ρ_G = gas density, lb/ft^3
V_g = V = superficial gas velocity, ft/sec or m/sec
Note: 1 in. water/ft = 1.87 mm Hg/ft = 6.15 mm Hg/m

(14-51)

where

14.25.1. Packing Efficiency/Performance for IMTP Packing [96]

This is usually expressed as HETP, and for the IMTP packing when a high performance (uniform) liquid distributor is used in a column, the HETP is independent of the tower diameter and packing depth. Norton [83] has developed a concept for evaluating HETP for fluids

of the types (a) nonaqueous, (b) nonreacting and non-ionizing, and (c) low relative volatility (less than 3) [83], as follows:

System base HETP of IMTP packing for distillation and reboiled stripping:

(14-52)

(14-53)

expressed as millimeters or inches, depending on the constants used for A and B in the previous equations. If $\sigma > 27$ use, $\sigma = 27$, dynes/cm; μ = liquid viscosity, cP. Use the values of A and B as follows [Reference 83 by permission]:

Values of A and B for Equations 14-52 and 14-53					
IMTP Size	**No. 15**	**No. 25**	**No. 40**	**No. 50**	**No. 70**
A-mm	271.3	331.4	400.7	524.9	759.9
A-in	10.68	13.05	15.78	20.67	29.92
B-mm	296.7	366.1	438.3	579.9	831.2
B-in	11.68	14.41	17.25	22.83	32.72

For commercial towers with good liquid/vapor distribution, Norton [96] recommends that standard designs use HETP values that are 13% above system base HETP. If the system under consideration does not meet the physical properties limit, either use a conservative estimate or use actual plant or published data. For comparison of HETP values for selected packings, see Strigle and Rukovena [94], Figure 14-28.

Figure 14-28

Packing comparison at design point for Intalox® ceramic saddles, metal pall rings, and metal Intalox®.

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Example 14-2, Hydrocarbon Stripper Design (Figure 14-29)

Design a packed tower splitter for a light hydrocarbon plant. The conditions of operation, as determined from material balance are:

Figure 14-29

LHC plant stripper using packed tower; for Example 14-2.

Operate at 430 psig, feed from ethylene tower at purification unit.

Feed:

Volume:	55% ethylene
	45% ethane

Specifications to be:

Overhead:	90% ethylene
Bottoms:	95% ethane
Feed:	13,000 lb/day ethylene
	11,000 lb/day ethane

Determine the feed temperature from ethylene tower.

Solution

Bottoms: 25°F, Boiling point feed

Bubble point of bottoms: At 445 psia

Composition	**x_B**	**K @ 47°F**	**Kx**	**K @ 45F**	**Kx**
Ethylene	0.05	1.36	0.068	1.35	0.067
Ethane	0.95	1.0	0.95	0.97	0.921
	1.00		1.018		0.988

Therefore, bubble point = 46°F

Dew point on overhead: at 445 psia

Composition	**y_{OH}**	**K@15°F**	**y/K**
Ethylene	0.90	1.05	0.857
Ethane	0.10	0.75	0.138
	1.00		0.995

Dew point on overhead; 15°F

Bubble point on overhead: Try 10°F at 445 psia

Composition	x_D	K	Kx
Ethylene	0.90	1.00	0.90
Ethane	0.10	0.80	0.08
	1.00		0.98

Bubble point > 10°F < 15°F ≅ 12°F

Minimum Trays at Total Reflux

Fenske Equation:

Minimum Reflux

Gilliland Plot:

Theoretical Plates vs. Reflux

Gilliland Plot

(L/D)	Theoretical Plates
3.25	Infinity

3.25	Infinity
4.0	28.4
5.0	22.4
6.0	20.0
7.0	18.6
8.0	17.8
Infinity	14.2

Figure 14-29 shows the schematic diagram of loading and Figure 14-30 shows the plot for this example respectively.

Figure 14-30

Gilliland plot for Example 14-2.

From the material balance using (L/D) = 6.0; theoretical trays = 20.

(a) Feed:	24,000 lb/day
(b) Distillate product:	13,900 lb/day
(c) Reflux into column:	83,400 lb/day @ 6:1 reflux, L/D
(d) Total gross overhead:	97,300 lb/day
(e) Total bottoms out:	10,100 lb/day

Top of column

Liquid viscosity = 0.07 cP

From Figure 14-21C:

Read:

For average loading condition, read ordinate = 0.030

For average flooding: (from chart) using 1-in. metal Pall rings with F = 56.

Area required = 0.4559/0.667 = 0.6835 ft^2 (= 98.4 in^2)

= 11.19 in.

Bottom of column:

Using packing factor, F = 56 for 1-in. Pall metals rings

Reading chart:

, flooding, for reference

Loading velocity, $V_{loading}$ = 1.5425/2.47

= 0.6245 ft/sec

Volume flow rate = 97,300/(24) (3,600) (2.47) = 0.4559 ft^3/sec

Area required = 0.4559/0.6245 = 0.7299 ft^2 (= 105.1 in^2)

= 0.964 ft. (11.56 in.)

Use 16-in. diameter, See Appendix A-18.

It is better to use smaller packing (less than 1 in.) in this diameter column (16-in. I.D.), such as ⅝ in. if recalculated above. Packing of ½ in. or ⅜ in. are better in this size column; however the effects of packing factor should be calculated if changed from ⅝ in. packing.

HETP: Use HETP = 18 in. based on on-site column data and manufacturer's confirmation that for ⅝ in. Pall rings in this system, the 18 in. HETP should perform satisfactorily. Note: For each design verify expected HETP through the manufacturer.

Redistribute the liquid every 10 ft and allow 2 ft additional for redistribution = 36 ft. Then use 4-10 ft sections of packing. For 20 theoretical plates, total performance packed height 20 (18 in./12 in.) = 30 ft. Allow for loss of equilibrium at (1) reflux entrance = 1 HETP and (2) feed entrance = 2 HETP and (3) redistribution (2) = 2 HETP; totals 5 HETP.

Total packing to install = 30 ft + 7.5 ft = 37.5 ft; Use 40 ft Feed point, based on Kirkbride (See Chapter 10):

(14-54)

Allow 2 ft space between support of top section and redistributor of section 2 (from top).

Estimated pressure drop of loading from Figure 9-21C: = 0.40 in water/ft of packing. For total packed height, $\Delta P \cong (0.40)(16.5 + 24) = 16.2$ in. water total (packing only).

Preliminary evaluation of condenser requirements: Condense: 97,300 lb/day

For approximate estimate: assume overall heat transfer coefficient,

Recommend kettle type condenser (boiling propylene (see chapter on Heat Transfer, Volume 3).

Nutter Ring [97]

Nutter offers an improved high performance random packing identified as Nutter Ring™; see Figure 14-6K. To achieve the best performance from any random packing, the liquid distributor must be well-balanced, and the distribution points of the discharging liquid to the packing must be uniformly distributed; see earlier discussion on this topic.

The manufacturer's key performance descriptions and claims are:

1. Comparative tests at Fractionation Research, Inc. (FRI) [154] showed the No.2 Nutter ring™ to have greater usable efficiency-capacity profile, with substantially lower pressure drop than is achieved with slotted rings. Usable capacity is comparable to 3½-in. Pall rings (see Figure 14-31AFigure 14-31B and Figure 14-31C).

Figure 14-31A Comparison of HETP for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan

distributor.

Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc.; all rights reserved.

Figure 14-31B

Comparison of pressure drop for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan distributor.

Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc.; all rights reserved.

Figure 14-31C

Comparison of P/Eo for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan distributor.

Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc.; all rights reserved.

2. Ring form provides multiple circular crimped strips to encourage liquid surface renewal.
3. Perforated central trough enhances lateral liquid spreading and effectively wets the outside surface facing vapor flow.

4. Heavily ribbed main element gives a high strength-to-weight ratio.
5. A pair of tapered slots and hoops provide maximum randomness with minimal nesting.
6. Efficiency enhanced by item (2) above, and the No.2 Nutter Ring is better than 2-in. Pall rings [155, 157].
7. Superior surface utilization in mass and heat transfer, allowing shorter packed bed heights. Turndown performance is superior to 2-in. and 3½-in. Pall rings.
8. Strength to weight ratio allows bed heights of up to 50 ft.
9. Lower pressure drop and greater usable capacity allow use of smaller diameter columns. Pressure drop per stage is 30–50% less than either size Pall ring.
10. Reproducible performance assured through uniform randomness [156].
11. All rings in sizes No. 0.7 to No. 3 diameter are proportional on all dimensions for accuracy in scale-up of pressure drop, capacity, and efficiency. See Figure 14-6K.
12. Plastic Nutter Rings™ are rigid and energy efficient, and permit applications to produce pressure drops per theoretical stage and bed heights that are not attainable with other random particle packings.

Capacity Correlation [98]

This design presentation is proprietary to Nutter Engineering and is definitely more accurate than applying the Nutter Ring™ to the generalized pressure drop correlations (Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21E and Figure 14-21F). Figure 14-21G and Figure 14-21I do not apply because they are proprietary to another manufacturer. The procedures to follow supersede the equations in Nutter and Perry [99, 98], and are used by permission of Nutter Engineering, a Harsco Corp.

Determine % useful capacity; assume a column diameter, or calculate using an existing column under study. Usable capacity is defined as the maximum loading condition where efficiency does not deteriorate significantly from that achieved over a lower range of loadings [99]. For the Nutter ring, the limiting pressure drop for usable capacity is 1 in. of hot liquid for low pressure systems. Designs should not exceed this value. This has been shown to hold for C_6-C_7 system at 5 psia and for hydrocarbons at atmospheric to 24 psia (see Figure 14-31A and Figure 14-31B) [98].

(14-55)

where

C_s = vapor rate = $V_s\,[\rho_v/(\rho_l - \rho_v)]^{0.5}$, ft/sec (14-56)

ρ_v = vapor/gas phase density, lb/ft^3

ρ_l = liquid phase density, lb/ft^3

V_s = vapor superficial velocity, ft/sec

C_2 = wet pressure drop intercept coefficient

(14-57)

(14-58)

(14-59)

(14-60)

(14-61)

Table 14-35 Nutter Ring Hydraulic Coefficients

Used by permission of Nutter Engineering, a Harsco Corp.

Nutter ring(TM) size	C_1	X_1	X_2	X_3	X_4
0.7	0.141	0.735	−0.00646	1.0	0.05

1	0.095	0.870	−0.00562	1.10	0.08
1.5	0.070	0.969	−0.00230	1.12	0.08
2	0.059	1.000	0.0	1.15	0.08
2.5	0.051	1.0394	0.000653	1.15	0.08
3	0.037	1.124	0.002076	1.15	0.10

(14-62)

σ = liquid surface tension, dynes/cm

(14-63)

not applicable when $((\rho_l - \rho_v)/\rho_v)^{0.5} > 4.5$

Pressure drop: dry bed

(14-64)

where C_1 = coefficient from Table 14-35.

Operating pressure drop:

(14-65)

where

(14-66)

C_3 = wet pressure drop slope coefficient

(14-67)

Z_{Q1} = liquid rate factor for C_3, in. water/ft
X_3, X_4 = constants from Table 14-35
ΔP_d = dry bed pressure drop, in. water/ft
ΔP = operating pressure drop, in. liquid/ft
e = base of natural logarithms
X_1, X_2 = curve fit coefficients for C_2, Table 14-35
X_3, X_4 = curve fit coefficients for C_3, Table 14-35
μ = viscosity, centipoises, cP
Subscripts
g,v = gas or vapor phase
l = liquid
s = based on tower crosssectional area

Figure 14-32A and Figure 14-32B [98] illustrate the correlation of wet pressure drop with system vapor rate at various liquid rates for No.2 Nutter rings; however, other available data indicate that other sizes of Nutter rings, Pall rings, and selected other packing shapes correlate in the same manner.

Figure 14-32A Correlation of No. 2 Nutter Rings™ superficial capacity vs. wet pressure drop for 4 data sets and 3 separate tests. Note the 10:1 pressure drop range.

Reproduced by permission from Nutter, D. E. and Perry, D., presented at New Orleans, La. meeting of American Institute of Chemical Engineers, March (1988), and by special permission of

Fractionation Research, Inc.; all rights reserved.

Figure 14-32B

Correlation of No. 2 Nutter Ring™ for 7 liquid rates versus wet pressure drop.

Reproduced by permission from Nutter, D. E. and Perry, D., presented at New Orleans, La. meeting of American Institute of Chemical Engineers, March (1988), and by special permission of Fractional Research, Inc.; all rights reserved.

Figure 14-33A and Figure 14-33B illustrate the fit of data measured by FRI on a commercial size column, for hydrocarbon systems, using No. 2.5 Nutter rings at three different pressures, and comparing the latest Nutter proprietary correlation previously presented. When considering pressure drop models based only on water, hydrocarbon system capacity can be significantly overstated. For Nutter random ring packings, the pressure drop/capacity models fit the data within ±10% over the range of commercial interest, i.e., 0.1 to 1.0 in. water/ft of packing. Pressure drop values for design operation should not exceed 1.0 in. water/ft. This is generally 80% of flood capacity or 90% of useful capacity.

Figure 14-33A Comparison of hydrocarbon systems fit to Nutter Correlation at 165 psia (No. 2.5 Nutter Ring™).

Used by permission of Nutter Engineering Co., Harsco Corp. and by special permission of Fractionation Research, Inc.; all rights reserved.

Figure 14-33B

Comparison of hydrocarbon systems fit to Nutter Correlation at 300 psia (No. 2.5 Nutter Ring™).

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Tests by FRI and Nutter [132] emphasize that distribution of liquid must be uniform, and at minimum values, to achieve good HETP values over a range of system pressures for hydrocarbon distillation. Glitsch performance data for their Cascade Mini-Ring® (CMR) are shown in Figure 14-34AFigure 14-34B and Figure 14-34C a longside other published data and pressure drops for comparison with Pall rings and sieve trays.

Figure 14-34A Efficiency versus C-factor for various metal packings. Iso-octane/toluene, 740 mm Hg, reflux ratio 14:1, 15-in. I.D. column, 10-ft bed height. Not Glitsch test data. Note: CMR = Cascade Miniring™.

Used by permission of Glitsch, Inc., Bull. 345.

Figure 14-34B

Efficiency versus C-factor for metal rings.

Data from Billet (Symposium Series 32, Institution of Chemical Engineers), London (1969). Used by permission of Glitsch, Inc., Bull. 345.

Figure 14-34C

Pressure drop versus C-factor for metal rings and sieve trays. Operating data from ethylbenzene/xylene service. 50 mm Hg top pressure. Naarden International test data.

Used by permission of Glitsch, Inc., Bull. 345.

Generally, it is not recommended to specify any packed section in a random packed tower to be greater than 20 ft in height. However, some packing manufacturers state that their packings will physically sustain greater heights and continue to produce good HETP values. This is done by maintaining good uniform liquid flow internally, and liquid/vapor internal contact by preventing significant channeling that would reduce the contact efficiency, This statement is based on factors that include creating the distribution of liquid, redistribution of liquid, gas or vapor channeling, and process surging, plus many other situations unique to the process conditions. Structured packing heights should be determined by the manufacturer for the design conditions.

When trying to “balance” the several packed section heights required for the process, it is completely acceptable to vary the heights of each section to fit such requirements as location of return of reflux, multiple feed positions, and safety factors. Thus one section may be

20 ft, another 17 ft, and another 25 ft as long as the process function has been thought through. The locations of the “breaks” in the packing sections must not interrupt control functions by forcing the location of the temperature sensor Because of sensing location. Into a non-optimal this, along with other reasons, it is good to prepare a profile of the column and spot the controls and “interruptions” to packing continuity.

One issue is that at least 3 feed injection nozzles need to be placed in the column during packing (installation), and then piped-up externally to the column's best performance determine conditions, and to allow for the monitoring of feed composition. In summary, everything that can be done mechanically, including piping, should be built into the column arrangement, in order to be able to accommodate future changes in process requirements, errors or lack of agreement between calculations and actual performance in operation.

14.25.2. Dumped Packing: Gas–Liquid System Below Loading

Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21EFigure 14-21FFigure 14-21G and Figure 14-21H can be used to determine performance for new designs, or for existing operating columns.

14.25.3. Dumped Packing: Loading and Flooding Regions, General Design Correlations

Figure 14-21BFigure 14-21F and Figure 14-21H may be used for any system to obtain a good estimate of pressure drop for practically any random packing material. The relative liquid can position, and the approximate pressure drops per foot of packing depth may be read as parameters. It is important to recognize that the load upper limit, line A, is essentially coincident with the flooding condition. It is also apparent that the relative relation of the operating point to the flooding and loading conditions is quite different at the extreme right of the figure for large values of the abscissa than for low values to the left.

The rearranged form of the same Sherwood equation [61, 62] allows the curve for flooding of dumped packings to be presented so as to facilitate calculation of the flooding gas rate, G_f, corresponding to a given liquid flow L [81], Figure 14-21A.

The packing factors to use with Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21EFigure 14-21F and Figure 14-21H are given in Table 14-26ATable 14-26BTable 14-26CTable 14-26DTable 14-26ETable 14-26F and Table 14-26G.

Ward and Sommerfeld [130] present an equation based on the curves shown in Figure 14-21C and Figure 14-21D and referenced to Eckert [125] and Leva [43], for calculating the gas and liquid flooding rates. There have been numerous other equations developed for this purpose, but many are too awkward for general use. The proposed equations have been tested by the authors.

Gas flow rate at flooding:

(14-68)

liquid flow rate, L, at flooding:

(14-69)

where

A =

B =

A = correlating parameter

B = correlating parameter

F = packing factor, 1/ft, or, $(ft)^{-1}$

G = gas flow rate, lb/(hr) (ft^2)
g_c = Newton's Law, proportionality factor
= (4.17×10^8), $(ft\text{-}lb_m)/(lb_f\text{-}hr^2)$
L = liquid flow rate, lb/(hr-ft^2)
X = abscissa in the generalized pressure drop correlation,
= $(L/G)\ (\rho_G/\rho_L)^{0.5}$
Y = ordinate in the generalized pressure drop correlation

μ = viscosity, centipoise
ρ = density, lb/ft^3
Ψ = ratio of water density to liquid density
Subscripts
G = gas phase
L = liquid phase

The solution of the previous equations requires careful attention to the sequence of the arithmetic. One difficult requirement is the need to establish the L or G in 1b/hr/ft^2 of tower crosssection, which requires a value for the tower diameter to be assured. The equations are quite sensitive to the values of A and B.

14.25.4. Dumped Packing: Pressure Drop at Flooding

As a comparison or alternative procedure, the pressure drop at the flooding point, as indicated by the upper break in the pressure drop curve, can be estimated from Table 14-36 and Figure 14-34D, for rings and saddles [81]. The values in the table multiplied by the correction ratio give the pressure drop for the liquid in question, expressed as inches of water.

Table 14-36 Pressure Drop at Upper Break Point (Flood) With Water As the Flowing Liquid [81].
By permission, F. A. Zenz, *Chem. Eng.*, Aug., 176 (1953), Ref. 81.

Raschig rings, in.	ΔP_f, in. H_2O/ft of packed bed	Berl saddles, in.	ΔP_f, in. H_2O/ft of packed bed
2	2.5	1½	2.2
1½	2.5	1	2.5
1¼	4.0	¾	2.5
1 (ribbed)	3.0	½	2.0
1	4.0	¼	1.25
¾	3.0		
⅝	2.5		
½	3.5		
⅜	4.0		
¼	4.0		

Figure 14-34D

Pressure drop correlation at flood point for use with Table 14-33.

Used by permission of Zenz, F. A., *Chemical Engineering*, Aug. (1953) p. 176; all rights reserved.

14.25.5. Dumped Packing: Pressure Drop Below and at flood Point, liquid Continuous Range

Liquid Continuous Range

For a particular liquid-gas system and tower packing, performance indicates a region where the liquid phase becomes continuous and the gas phase discontinuous. This is obviously at relatively high liquid rates, but not beyond the range of satisfactory performance for the equipment. This region is characterized by proportionally higher pressure drops than the gas-continuous region, and the existence of a critical liquid rate as this pressure drop deviation occurs.

Referring to Figure 14-20, the curve for L′-12,500 shows the beginning of the “move to the left”, swinging away from the uniform slope of the curves for lower L' values.

This probably is not the L_c value for the system. The study of Zenz suggests that the critical liquid rate, L_c, is the minimum liquid rate that completely wets the packing, thus having essentially all packing surface available for gas contact. Rates above this value should be determined by allowable pressure drop and the limitation that the tower often begins to approach flooding conditions more rapidly than in the gas-continuous region. Figure 14-35 correlates this L_c for Raschig rings and Berl saddles, as a function of liquid viscosity.

Figure 14-35

Values of liquid rate when the system becomes liquid continuous, L_c.

Used by permission of Zenz, F. A., *Chemical Engineering*, Aug. (1953) p. 176; all rights reserved.

More work is needed to fully understand this feature of tower performance, and to extend the existing information to other common packings. Determination of L_c from the figures will indicate whether the tower is operating under gas-continuous (values of L lower than L_c) or liquid-continuous (values equal to or larger than L_c). The approximate degree of wetting of the packing can be evaluated as the ratio of L/L_c [73]. The pressure drop is evaluated using Figure 14-21BFigure 14-21CFigure 14-21DFigure 14-21EFigure 14-21F and Figure 14-21H to determine the flooding liquid rate, L_f. The ratio of L_f

to actual L is calculate by reading Figure 14-36 to obtain ΔP actual/ ΔP_f. Thus, ΔP actual is the ratio value mulitiplied by ΔP_f, as calculated using Figure 14-34D and Table 14-36Table 14-24 and Table 14-24A.

Figure 14-36

Pressure drop correlation at flooding.

Used by permission of Zenz, F A., *Chemical Engineering*, Aug. (1953) p. 181.

14.25.6. Pressure Drop Across Packing Supports and Redistribution Plates

Little useful correlated information on pressure drop across packing supports and redistribution plates is available. Some order of magnitude guide data are given in Figure 14-37Figure 14-38Figure 14-39Figure 14-40 and Figure 14-41.

Figure 14-37

Comparison effect of pressure drop across support plate and bed of 1½ in. Intalox® saddles.

Used by permission of Leva, M., Lucas, J. M., and Frahme, H. H., *Ind. Eng. Chem*. V. 46, No. 6 (1954); all rights reserved.

Figure 14-38

Effect of choice of support plate on flooding rate.

Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Equipment Corp.).

Figure 14-39

Pressure drop through fused ceramic Raschig ring plate with plate covered with 1 in. of 1-in.-size Intalox®saddles.

Used by permission of Leva, M., Lucas, J. M., and Frahme, H. H., Ind. *Eng. Chem.*, V. 46, No. 6 (1954); all rights reserved.

Figure 14-40

Pressure drop through plain ceramic plate.

Used by permission of Leva, M., Lucas, J. M., and Frahme, H. H., *Ind. Eng. Chem.*, V. 46, No. 6 (1954); all rights reserved.

Figure 14-41

Flooding on perforated support plate.

Used by permission of Lerner, B. J. and Grove, C. S., Jr., *Ind. Eng. Chem.*, V. 43, No. 1 (1951), p. 216; all rights reserved.

Because these data are peculiar to the supports studied, they can serve only as a good estimate for other situations. It is important to remember that results obtained with a bare plate, and one holding a layer of packing can be quite different, the latter being the more realistic condition.

The only available test data [44] indicate that the plain flat plate (26–45% free area) has a decidedly detrimental effect on the allowable flooding conditions of the tower; whereas the wire screen, weir-type, and fused Raschig ring designs have very little effect when using Intalox saddles for packing.

In general the dumped saddle type packing should give less blocking to support openings than ring type.

Example 14-3: Evaluation of Tower Condition and Pressure Drop

Check the design of a 4-ft, 6-in. I.D. tower packed with 45 ft of 1-in. × -in. thick steel Raschig rings if the service requires a liquid rate of 2,250 lb/hr (ft^2) of 10% caustic solution (sp. gr. = 1.22) and 4,540 lb/hr (ft^2) at 110°F air containing CO_2 to be scrubbed at 365 psia.

1. Determine the operating range for the tower by referencing to Figure 14-21B and Figure 14-21C. Use Figure 14-21C for this example.

Solution

Table 14-26A note range.

2. Locating 0.0756 and 0.0563 on Figure 14-21C reading the intersection indicates a condition in the lower loading region, and a pressure drop of approximately 0.60 in. water/ft of packing.
3. Expected pressure drop through bed:

4. This bed should be in three sections (two might be acceptable under some circumstances, or if different packing were used) thereby requiring two intermediate combined packing supports and redistribution plates and one bottom support plate.
Estimate pressure drop per redistribution set-up or support

For two redistributors plus one support plate:

5. Total estimated pressure drop through packed portion of tower:

Bed ΔP	= 27.0

Bed ΔP	= 27.0
Internals	= 3.0
	30.0 in. water

Say, 30 to 35 in. water (this should be 30.0 ± 15–20%)

6. Estimated percent of flooding

Reading the abscissa of Figure 14-21C at 0.0756 for ordinate at flooding line for dumped packing: = 0.13
Then, actual value of 0.0563 is:

Note that this rather low value will usually occur when operating at the low L/G values due to the greater operating spread between flooding and loading (see Figure 14-21C).
7. Estimated percent of loading (average)
Reading abscissa at 0.0756 for ordinate of line B gives ordinate = 0.078

This is an acceptable value and should not be exceeded except in known systems. It is preferable to operate reasonably close to the loading condition for maximum contact efficiency.

As an alternative consideration, assume various pressure drops/foot of packing (same) and determine the effect on calculated column diameter. Use the same input information as original stated conditions, then:

Assumed Pressure Drop, in. H_2O/ft	**Calculated Column Diameter, ft**
0.25	5.22
0.50	4.61
0.75	4.35

Example 14-4: Alternate Evaluation of Tower Condition and Pressure Drop

Using Example 14-3 as reference, the tower will be examined for critical liquid rate, L_c, using Figure 14-35.

reading: L_c = 16,000 lb/(hr) (ft^2)

Because actual L′ = 2250 is less than 16,000; this tower operates in the gas continuous region.

Pressure drop at flooding from Table 14-36 and Figure 14-34D

ΔP_f = 4 in. water/ft (not exact figure, because table is for ceramic rings)

Correction = 0.94

Then actual expected pressure drop at flooding:

For 45 ft of packing

Comparison with Figure 14-21C gives 3 in. water/ft (parameter) or a total of (3) (45) = 135 in. water. Neither of these values represents a condition (flooding) that should be considered for tower operation, except for very well known systems. Distillation operations sometimes operate above flooding, but other types of contacting normally require operations in the loading region (or below) for stable

performance.

Example 14-5: Change of Performance with Change in Packing in Existing Tower

A tower is packed with l-in. ceramic Raschig rings. It presently floods while drying a product at a production feed rate of 1,800,000 lbs/month with 0.25 mol% being water. Flooding does not start at the bottom, but at some intermediate point up the tower. What can be done to eliminate the flooding? Is it possible to increase production rate to 2,000,000 lbs/month?

Solution

1. Examine packing characteristics.

Size	a/ε^3	Surface area, ft^2/ft^3
1 in. Rashig Rings	158	58
1 in. Intalox	124	78
1 in. Berl Saddles	125	76
1 in. Pall Ring*	45	66.3
1½ in. Intalox	69	59.5
1½ in. Berl Saddle	79	44
*metal, all others ceramic		

2. Percent of flooding for various packings, holding tower flow rates (including reflux) constant.

Refer to Flooding, Loading and Pressure Drop Chart, Figure 14-21D.

remains constant for same separation at increased production rate.

increases as G^2 at increased production rate for a fixed a/ε^3.

Packing	Percent of Flooding at 2,000,000 lb/mo. rate referred to flooding at 1 in. ceramic Raschig rings
1 in. R.R.	100%
1 in. Intalox	
1 in. Pall Rings (metal)	
1½ in. Intalox	
1½ in. Berl Saddle	

The flooding of the packing is a direct function of the a/ε^3, therefore it is valid at constant separation to examine the performance as shown. The metal Pall rings appear to allow for a considerable increase in capacity. In fact the condition at 35.2% of flood might not be good from a contact efficiency standpoint.

3. Selection

The 1½ -in. Intalox or Berl ceramic saddles would be the preferred choice because: (1) the flooding point is sufficiently low and yet probably not too far from the load point (only flood data available, but would estimate 70–85% of load); (2) the surface area per cubic foot is essentially the same as for the existing 1-inch Raschig rings. By reference to the effective interfacial area graphs, and by using the Berl saddle data instead of Intalox as an estimate because it is not available, the separation performance would be expected to be essentially identical to the existing tower (for the lntalox); (3) the production rate can be increased; (4) the flooding can be removed from present operations by using the 1½ -in. Intalox or any of the other packings, but note that the a/ε^3 of the Berl saddles is not as high as the l-in. R. R. This might mean less efficient contact, requiring more packing. However, Figure 14-42 note that for a given L'/G', the Berl saddles are 15–30% better wetted in any given packing volume. From Figure 14-43 and Figure 14-44, the comparative effective areas, (a), indicate that the Berl saddles have about the same area/ft^3 as the existing rings. Therefore, it appears that the flooding can be stopped (lower a/ε^3) and yet the contact can be as good, or maybe a little better than it was originally.

Figure 14-42

Total hold-up for ceramic tower packings: air-water.

Used by permission of Shulman, H. L., Ullrich, C. F., and Wells, H.; *A.Ch.E. Jour.*, V. 1, No. 2 (1955) p. 247; all rights reserved.

Figure 14-43

Fraction packing wetted.

Used by permission of Shulman, H. L., Ullich, C. F., Proulx, A. Z., and Zimmerman, J. 0., *A.I.Ch.E. Jour.*, V. 1, No. 2 (1955) p. 253; all rights reserved.

Figure 14-44

Gas-liquid hold-up data for ceramic rings and saddles.

Used by permission of Leva, M. *Tower Packings and Packed Tower Design*, 2nd ed., U.S. Stoneware Co. (now, Norton Chemical Process Equipment Corp.) (1953).

4. Check support grid voidage

The packing support consists of a floor grating material with bar openings spaced to give 57.5% free void area of crosssection.
To avoid the possibility of local flooding at the support, it would be well to place a heavy hardware cloth over this grid, to keep the saddles from nesting at the first layer in the opening slots. It is preferable to have them resting on a surface that cannot be easily blanked.
Select a l-in. × l-in. (center line) × 0.063-in. wire cloth with voidage of 87.8%.
Combined voidage, support grid plus cloth:
= (0.575) (0.878) = 50.5%
This should be satisfactory, but a value much lower than this could not be tolerated.
Dump Intalox (or Berl) saddles into tower while tower is filled with water.

Example 14-6: Stacked Packing Pressure Drop

Consider the problem of drying air with sulfuric acid, Example 14-7. If the mass transfer relations are evaluated, a reasonably good estimate of the new packed height can be determined for using 2-in. stacked ceramic Raschig rings in place of Intalox saddles. For the present, assume the new required stacked height is 25 ft. Although the pressure drop per foot will be less, and the flooding point higher, than the dumped packing, it is inadvisable to use a smaller diameter tower, because of the high superficial gas velocity.

Using the 15-in. I.D. ceramic tower, the expected pressure drop will be:

From Reference 40 for 2-in. ceramic stacked Raschig rings:

$\alpha = 0.06$

$\beta = 0.012$

Liquid range checks satisfactory

$G = 0.593$ lb/(sec) (ft^2)

Leva [40] has shown that for liquids other than water, L must be corrected by the ratio of the density of water to that of the fluid in the system.

Total tower drop:

Packing	= (0.00258) (25) = 0.064 in. water (*approximate*)
Support (estimated)	= 1.5 in. water
Total (approximate)	= 1.56 in. water

Note that the weight of liquid will be greater in this arrangement at flooding, and the operating hold-up will be almost the same as the

dumped Intalox. The total weight of packing will be approximately 50% greater than if the same 2-in. rings had been dumped in place. Two-inch rings are not usually stacked. In this small tower made up of 3-ft ceramic sections, the stacking is not too difficult a job if there are conditions which justify the extra effort and expense.

14.26. Liquid Hold-up

Liquid hold up in a tower represents the liquid held in the void spaces of the packing during operating conditions. At flooding, essentially all of the voids are filled with liquid.

Usually low hold-up is desired, but same is necessary for efficient tower operation. The weight of liquid held in the packing must be considered when determining the support loads at the bottom of the packing, as well as the tower itself. The higher the hold-up for any particular packing, the greater will be the gas pressure drop, and the longer the tower drainage time when it is shut down. Smaller size packing tends to have greater hold-up than larger packing.

Figure 14-44 presents water hold-up data that are correlated by [40]:

(14-70)

where

h_{tw} = water hold-up, ft^3 liquid/ft^3 volume tower

L' = Liquid rate, lb/(hr) (ft^2)

d_p = Equivalent spherical packing diameter, inches (diameter of packing equivalent to sphere of same surface area as the packing piece). Values given for some packings with physical data Table 14-1Table 14-2Table 14-3Table 14-4Table 14-5Table 14-6ATable 14-6BTable 14-7Table 14-8Table 14-9Table 14-10Table 14-11Table 14-12Table 14-13 and Table 14-14. Area of

sphere = π (diameter)^2.

For liquids other than water [36, 40]:

(14-71)

where

h_l = 1iquid hold-up, ft^3/ft^3 packed tower volume

μ_L = liquid viscosity, centipoise

ρ_L = liquid density, lb/ft^3

σ = surface tension, dynes/cm

Values of exponent n are given in Figure 14-45.

Figure 14-45

Liquid hold-up variation of surface tension exponent with liquid rate.

Reproduced by permission of the American Institute of Chemical Engineers; Jesser, B. W., and Elgin, J. C. *Trans. A.I.Ch.E*. V. 39, No. 3 (1943) p. 295; all rights reserved.

Total liquid hold-up in packed bed, h_t = static hold-up, h_s, plus operating hold-up, h_o [64, 66].

The static hold-up is independent of liquid and gas flow rates, since it represents the liquid held in the packing after a period of drainage time, usually until constant weight of material is received. This

requires approximately 1 hour for a 10-in. dia. × 36-in. packed height tower. Table 14-37 adequately summarizes the data.

Table 14-37 Static Hold-up in Random Packings

By permission of The American Institute of Chemical Engineers, Shulman, H. L. et al., *Chem. Engr. Jour.* Vol. 1, No. 2, 247 (1955) and ibid, p. 259 (1955), Ref. 64 and 66, all rights reserved.

Packing nominal size, in.	Static water hold-up, h_s, ft^3/ft^3 packing
Raschig rings (unglazed porcelain)	
½	0.0325
1	0.0150
1½	0.0089
2	0.0058
Berl saddles (unglazed porcelain)	
½	0.0317
1	0.0110
1½	0.0052
Raschig rings (carbon)	
1	0.0358
1½	0.0200
2	0.0150

Total hold-up, h_t, of water is represented, for Raschig rings and Berl saddles [66] as:

(14-72)

Constants are given in Table 14-38.

Table 14-38 Total Hold-up Constants

By permission of the American Institute of Chemical Engineers, Shulman, H. L. et al., *Chem. Engr. Jour.* Vol. 1, No. 2, 247 (1955) and ibid, p. 259 (1955), Ref. 64 and 66, all rights reserved.

Packing	α	γ	θ
Porcelain Raschig ring	2.25×10^{-5}	0.965	0.376
Carbon Raschig ring	7.90×10^{-5}	0.706	0.376
Porcelain Berl saddles	2.50×10^{-5}	0.965	0.376
Porcelain, in.	**Equivalent spheirical dia., ft**		
½ R. R.	0.0582		
1 R. R.	0.1167		
1 ½ R. R.	0.1740		
2 R. R.	0.238		
½ Berl saddle	0.0532		
1 Berl saddle	0.1050		
1½ Berl saddle	0.155		
Carbon, in.			
1 R. R.	0.1167		
1½ R. R.	0.178		
2 R. R.	0.235		

Figures 14-42A–F present the graphical interpretation of the total hold-up equation. These are more convenient to apply where the system fits (or nearly fits) the curves.

These data are valuable for determining the total weight of liquid held in the packing, and also the void fraction, in an operating column. ε is the void fraction of the dry packing minus the total hold-up, h_t.

14.26.1. Correction Factors for Liquids Other Than Water

In order to use the data in systems handling liquids other than water, correction equations and charts are used [66]. The charts are more convenient to use, and are presented in Figures 14-46 A–D. First, determine the total or static holdups for water at 20°C; second, determine separately the correction for viscosity, density, and surface tension; third, multiply the water hold-up by each of the corrections to obtain hold-up for liquid of the specific system.

Figure 14-46

Physical property corrections for liquid hold-up for ceramic packing and carbon packing (as noted). Reproduced by permission of *A.I.Ch.E. Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all rights reserved. **A.** Factors to be applied to water static holdups to determine the effect of surface tension. Used by permission of American Institute of Chemical Engineers, *A.I.Ch.E. Jour.*, Shulman, H. L., Wells, N., UIIrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all rights reserved. **B.** Factors to be applied to water operating holdups to determine effect of surface tension. Reproduced by permission of the American Institute of Chemical Engineers, *A.I.Ch.E. Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all rights reserved. **C.** Factors to be applied to water holdups to determine effect of Iiquid density. Used by permission of the American Institute of Chemical Engineers, *A.I.Ch.E. Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all rights reserved. **D.** Factors to be applied to water holdups to estimate the effect of liquid viscosity.

Used by permission of the American Institute of Chemical Engineers, *A.I.Ch.E. Jour.*, Shulman, H. L., Wells, N., UIIrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all rights reserved.

A second hold-up correlation reported by T. Otake and K. Okada [55] represents a survey of a considerable amount of literature, and is applicable to aqueous and nonaqueous systems for Reynolds numbers in the range 10–20,000 [40].

(14-73)

14.27. Packing Wetted Area

Wetted packing area may be considerably different to the physical area of a packing. This is of particular importance when comparing the effectiveness of different packings. It is only recently that a coordinated set of data became available for wetted areas in Raschig ring and Berl saddle packing [65].

Figure 14-43 represents the water-air and ammonia-water data for Berl saddles [65, 67]:

(14-74)

and for Raschig rings by

(14-75)

where

f_a = fraction of total packing area, a_t, that is wetted

a' = wetted packing surface not necessarily same as effective interfacial surface area for contact, ft^2/ft^3

a_t = total packing surface, ft^2/ft^3
L' = superficial liquid rate, lb/hr (ft^2)
G' = superficial gas rate, lb/hr (ft^2)

The fractional wetted area immediately indicates the effectiveness of contact for the liquid system in the packing. This contact efficiency must be considered in some design problems.

14.28. Effective Interfacial Area

The effective interfacial area is used in mass transfer studies as an undivided part of individual and overall coefficients, when it is difficult to separate and determine the effective area. The work of Shulman *et al.* [65] presents a well organized evaluation of other work, in addition to their own. One of the difficulties in correlating tower packing performance lies in obtaining the correct values for the effective interfacial areas of the packing on which the actual absorption, desorption, chemical reaction, *etc.* occur. Figures 9-47 A–G present a correlation for water flow based on the ammonia-water data of Fellinger [27] and are valid for absorption work.

Figure 14-47

Effective interfacial areas for random ceramic tower packings. Note for gases or vapors other than air, use abscissa, G, as $G/(\rho gas/0.075)^{0.8}$. Based on the data of Fellinger [27].

Used by permission of the American Institute of Chemical Engineers; *A.I.Ch.E. Jour.*, Shulman, H. L., UIIrich, C. F., Proulx, A. Z., and Zimmerman, J. O., V. 1, No. 2, (1955) p. 253. All rights reserved.

There are differences between wetted and effective area as discussed by Shulman [65]:

1. wetted areas increase as packing size decreases.
2. effective area is smallest for the smallest packings.
3. the effective area seems to go through a maximum for l-inch packing, even though the larger packings have almost as much area.

This is better understood in terms of the hold-up data for these packings. For vaporization in packed beds of Raschig rings and Berl saddles [66];

(14-76)

(14-77)

where the subscripts *vap* and *abs* represent conditions of vaporization and absorption respectively, and subscript w represents a water system.

14.29. Entrainment From Packing Surface

There is not much data available for this subject; operational experience plus qualitative tests indicate that entrainment is negligible until the packing reaches the flooding condition. See discussion under distillation section.

Example 14-7: Operation at Low Rate, Liquid Hold-up

A sulfuric acid drying tower uses 98% acid for drying an incoming air stream. The pilot plant tests show that 15 ft of l-in. ceramic Intalox packing will do this job. The plant scale rates are:

Air = 500 scfm at 90°F and 2 psig

Acid = 6 gpm at 90°F and sp gr = 1.81

Determine (1) the tower diameter (2) pressure drop (3) liquid hold-up

Solution

Assume for first attempt:

Inside tower diameter = 12 in.

Reading Figure 14-21D at the calculated ordinate and abscissa, we obtain:

(a) An indicated operating condition above the upper loading limit

(b) A condition of high pressure drop, approximately 1.5 in./ft of packing height

(c) A situation too close to flooding, thus requiring a larger tower diameter

Second Attempt: Try 15-inch Dia. Ceramic Tower

Inspection of Figure 14-21BFigure 14-21C and Figure 14-21D shows that the increase in tower diameter does not affected the value of the abscissa. By changing the tower diameter to 15-in. crosssection area = 1.22 ft^2.

This indicates operation in the loading region. The expected pressure drop is 0.5 in. water/ft.

Total expected pressure drop:

Packing = (0.5) (15) = 7.5 in. water

Support = 1.5 in. (estimated from Figure 14-37Figure 14-38Figure 14-39 and Figure 14-40 for a 58% open grid).

Total drop = 9.0 in. water (approximate)

Superficial gas velocity through tower:

Entrainment

This velocity is slightly high, and so an entrainment knockout or separator should be installed in the air stream following the tower, or in the top of the tower itself.

Liquid hold-up in the tower:

For water, the hold-up would be, from Equation 14-70:

(14-70)

d_p = 0.68 (from the Table 14-7)

For sulfuric acid:

From Figure 14-46C, h_o/h_{ow} for density correction multiplier = 0.6

From Figure 14-46B, correction for surface tension = 1.0 (at 70 dynes/cm)

From Figure 14-46D, correction for viscosity = 1.1 (at 18 cP)

h_o, for acid = h_{ow} (0.6) (1.0) (1.1) = (0.0384) (0.66) = 0.0254 ft^3 acid/ft tower volume

For a packed volume of 15ft in a 15-in. I.D. tower, the total acid hold-up:

= [(15) (1.22)] (0.0254) (112.6 lb/ft^3)

Total hold-up = 52.3 lb acid

Weights

Weight of dry packing in tower:

Total weight on bottom support plate when operating (not flooded)

Some allowance should be made for surging or uneven operation.

The maximum expected weight of liquid would be at flooding conditions:

Using percent free gas space = 77.5

Volume of liquid space = (15) (1.22) (0.775) = 14.2 ft^3

Weight of acid in this space = (14.2) (112.6) = 1,600 lb

Maximum support load = 770 + 1,600 = 2,370 lb

This is the load that should be considered for the support design and selection. To allow for unusual conditions, specify support load = (1.1) (2,370) = 2,600-lb minimum.

14.30. Structured Packing

Structured packings in current use are composed of:

1. Wire-mesh weavings (Figure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EE and Figure 14-6FF).
2. Corrugated sheet, or crimped sheet (usually somewhat thin) (Figure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MM and Figure 14-6NN).
3. Grid-type; open, heavy (usually metal) bar-grid shapes stacked together (Figure 14-6OOFigure 14-6PPFigure 14-6QQFigure 14-6RRFigure 14-6SS and Figure 14-6TT).

The preferred process application for structured packings depends on the geometric arrangement of the components and:

1. High cost/ft^3 of space
2. Ease of plugging by solid particles.
3. Importance of uniform liquid distribution across the packing.
4. Whether the system is operating under pressure [153] or vacuum.

For most efficient performance, the fabricated sections of the structured packing usually are placed in a specific rotating pattern in the column to ensure uniform liquid flow, and vapor cross-mixing.

The materials of construction are usually stainless steel or other specific metals that will draw into wire, or crimp without cracking. The wire mesh types are also fabricated from some plastics such as Teflon®, polypropylene, etc.; however, the surface must be wettable by the liquid, or the efficiency will be poor. Performance data are needed to complete a good design.

Nutter [99] reports that most, if not all, structured packings (Figure 14-6LL and Figure 14-6MM) follow the linear relationship of vapor rate vs. pressure drop at fixed liquid rates as exhibited by random packings.

Structured packing ACS ST-100 (Figure 14-6DD) is reported by the manufacturer to be equivalent in all respects to the original Sulzer packing, *i.e.* extremely low pressure drop and high efficiency (low HETP) over a wide range of liquid load, and in high-vacuum applications. The design rating procedure used by this manufacturer is confidential (as is also the best and final procedure for most other structured and grid packings) and design application data should be submitted to them for recommendations.

Preliminary Sizing for ACS Industries Series Woven X/S Knitted Wire Mesh Structured Packing

The following is extracted by permission from ACS Bulletin B-129 (Apr. 1992). Typical HETP is 3 in. compared to ACS products ST-100 of 7 in. Approximate comparison data are: ΔP as low as 0.1 in. water/theoretical tray, turndown ratio about 20:1. Special series S construction recommended for smaller columns, and transverse Series X for larger ones. The suggested preliminary sizing procedure is:

1. Determine the number of theoretical plates/trays/stages using standard methods.
2. Calculate the minimum allowable superficial vapor velocity (flooding) for equal vapor and liquid loads, for the X-100 and S-

100 packings.

(14-78)

where

μ = liquid viscosity, cP

ρ_L = liquid density, lb/ft^3.

ρ_V = vapor density, lb/ft^3.

If liquid viscosity is less than 0.15 cP, use constant 0.53 in place of the $\mu^{0.33}$ term.

= calculated or estimated maximum superficial vapor velocity.

Calculate: L/G

L = total liquid load, lb/hr.

G = total vapor load, lb/hr.

Use Figure 14-48 with this L/G and find the load ratio correction factor. Multiply by this factor, to determine the corrected maximum superficial vapor velocity, , for X-100 and S-100 packings.

Figure 14-48

ACS maximum vapor velocity correction for L/G for woven/knitted wire mesh structured packing.

Used by permission of ACS Industries, Inc., Separation Technology Division, Bull. B-129 (1992).

Correct the maximum vapor velocity again for the packing type, if other than X-100 or S-100

(14-79)

where

= maximum vapor velocity for actual packing, ft/sec.

= maximum superficial vapor velocity for X-100 and S-100 packing.

SS_A = specific surface for the actual packing* (see Table 14-39ATable 14-39B and Table-39C).

Table 14-39A JPG Mechanical Characteristics of Common Standard Packing Styles (ACS Wire Mesh)
By permission of ACS Industries, Separations Technology Group, Bul. B-129 (1992).

ACS packing style	Number of strands	Strand diam., in.	Density, lb/ft^3 (stainless)	Pct. void vol.	Sp. surf., ft^2/ft^3
X-100, S-100*	12	0.0045	27.5	94.5	585
X-200*, S-200	8	0.0045	20.1	96.0	426

*For Series X, standard style is X-200 (8 strands)
For Series S, standard style is S-I00 (12 strands)

Table 14-39B Experimental HETP Values from Various Sources for Packings Equivalent to Stainless Steel X-200.
By permission of ACS Industries, Separations Technology Group, Bul. B-129 (1992).

System	Diam., in.	Absolute mm Hg	HETP, in.
Methylcyclohexane & toluene	1.25	760	1.25–2.00
	1.25	200	1.75–2.50
Methylcyclohexane & toluene	6	200–760	2–3
	18	200–760	4–6
Improved liquid distribution→	18	200–760	3.5
Methylcyclohexane & heptane	3	760	1.25–2
Ethylene dichloride & benzene	12	760	4–9
Orthodichlorobenzene & orthodiethylbenzene	1.25	50	1.25
	1.25	10	1.65
Phenol & orthocresol (no distributor used)	4	7–25	2.5–4
	4	85–100	2–3.5
	4	760	2–3.5
n-Hexanol & aniline	1.25	50–300	2.5–2.75
Nitrobenzene & aniline	1.25	5	3.5–4
n-Decanol & methyl naphthalene (no distributor used)	4	3	5
n-Decanol & methyl naphthalene	4	3	4
Methanol & water	1.25	760	9

Table-39C Rough HETP Correction Factor for Diameter.
By permission of ACS Industries, Separations Technology Group, Bul. B-129

(1992).

Column diameter	HETP proportional to:
Below 2 in.	1.0
2 to 6 in.	1.5
6 to 18 in.	2.0
Over 18 in.	2.5 to 3.0

SS_1 = specific surface for X-100 and S-100 packings (see Table 14-39).
PV_A = percent void volume for the actual packing∗, (see Table 14-39).
PV_1 = percent void volume for the X-100 and S-100 packings.

Calculate required column area and diameter using the maximum superficial vapor velocity , and the actual vapor load.

Convert the mass vapor load, G, (lb/hr) to the actual vapor volumetric flow, V, in ft^3/sec. Divide by to obtain the minimum crosssectional column area, A, ft^2.

Correct this area by dividing the fraction of capacity at which the designer intends for the column to operate. A value of 0.7 is recommended, unless some other consideration suggests a different percentage.

(14-80)

Final column inside net area = capacity factor.

Calculate the column net inside diameter from the area values. Round to the nearest practical commercial column diameter such as 12 in., 15 in., 18 in., 24 in., 30 in., 36 in., and above in increments of 6 in. Then recalculate the actual resulting vapor velocity.

(14-81A)

(14-81B)

Commercial size columns using Series X or Series S packing will operate at full efficiency with liquid loads as low as 5 gal/hr/ft^2 column area, and sometimes less. They can operate satisfactorily at liquid rates exceeding 2,650 gal/hr/ft^2. See Figure 14-49.

Figure 14-49

Estimated pressure drop referenced to percentage of maximum flood load for structured woven/knitted wire mesh X-100 and S-100. Referenced to liquid of 60 Ib/ft^3. To correct other liquids, multiply ΔP from figure by ratio of actual liquid density to 60.

Used by permission of ACS Industries, Inc., Separation Technology Division, Bull. B-129 (1992).

HETP for ACS Series X-200 Structured Packing:

1. If plant or pilot plant data are available, use them.
2. Table 14-39B presents references for packings equivalent to the X-200 for the systems and conditions noted. See Figure 14-50.

Figure 14-50

HETP and pressure drop data for a typical distillation system. Packing: equivalent to X-200 (8 strands), stainless steel. System: methylcyclohexane and toluene. Reflux Ratio: 100%. Column Diameter: 18 inches. Packed Height: 5 feet.

Used by permission: ACS Industries, Inc., Separation Technology Division, Bull. B-129 (1992).

3. Using an available HETP value for a specific Type X or S packing to convert to a column of different diameter, use the "rough" correction factors in Table 14-39C. These factors recognize that larger diameter columns have more uneven liquid distribution. When no HETP values or data are available, it is suggested to use the approximation developed for S-100 packing.

Estimating HETP.

(14-82)

where

HETP = height equivalent to a theoretical plate/tray/stage, ft.

μ = liquid viscosity, cP

R = liquid rate, gal/hr/ft^2

If $\mu < 0.30$ cP, use 0.52 in place of the $\mu^{0.55}$ term. To convert an available HETP value for a given system and column diameter to a different packing in the same series (X or S), assume HETP is inversely proportional to the specific surface (consult manufacturer for this detail).

(14-83)

where 1 and 2 = two different wire mesh packings.

For highly varying vapor loads through the column, pressure drop may be estimated for differences in latent heats from one point in the column to another; the pressure drop should be calculated for each foot of packing, or less adjusting the load. This particularly applies to vacuum columns or tall columns at high pressures. Maximum heights of a packed section in a column should be determined after consultation with the manufacturer.

Pressure Drop (Estimated):

Use Figure 14-49 to estimate column pressure per foot of packing height, referenced to liquid load as percentage of maximum flood load. For varying load, by perhaps ±20%, vacuum systems and varying latent heats in the column, calculate the pressure drop for each foot of packing or less; this requires adjusting the percentage load in each calculation.

Koch Kulzer Structured Packing [100]

This packing is woven wire mesh with the following characteristics as reported by the manufacturer/licensee [100]:

1. HETP of about 7 in. even at low liquid and vapor loads.

2. Efficiency essentially independent of column diameter.
3. Capacity and pressure drop characteristics superior to the best commercially available mass transfer devices.
4. Specific pressure drop of 0.2 mm Hg per theoretical tray.

This packing is a woven wire fabric of parallel corrugated elements and is completely self-wetting. It does not require pre-flooding to attain the wetted condition. This packing has a flat surface of approximately 150 ft^2/ft^3, and a free area > 90%.

In operation, the liquid flows downwards in a zig-zag pattern. See Figure 14-6EE.

Example 14-8: Koch-Sulzer Packing Tower Sizing (used by permission, Bulletin KS-1, Koch Engineering Co. Inc.)

The following typical design problem illustrates the calculation procedure for Koch Sulzer Packing. A comparison of a KS Packing design and a Flexi ring design will result in (for 2-in metal, Pall rings) top = 7.5-ft. D and bottom = 8.5-ft D. See Figure 14-51Figure 14-52Figure 14-53 and Figure 14-54.

Figure 14-51

Characteristics of Koch/Sulzer packing, Gas loading factor, F, versus HETP, pressure drop, and packing hold-up. Note: V_s = superficial gas velocity, ft/sec and ρv = vapor density, Ib/ft 3.

Used by permission of Koch Engineering Co., Inc., Bull. KS-1 and KS-2.

Figure 14-52

Koch/Sulzer packing. Design mass velocity (Ib/hr-ft^2) for HETP of 7 in. System cis-trans-decalin; molecular weight = 138.

Used by permission of Koch Engineering Co., Inc., Bull. KS-I.

Figure 14-53

Koch/Sulzer packing. Design mass velocities for systems other than cis-trans-decalin are obtained by dividing the values from Figure 14-52 by the "f" factor. For the following groups corrections should be applied to the number of carbons in the molecule:

Paraffins, olefins acids, aldehydes, ketones	0
Benzene ring, alcohols, phenol	−1.0
Saturated ring	−0.5
Nitro	+1.0
Ether	+1.5
Ester, secondary amine	+2.0
Chlorine	+3.0
Bromine	+8.0

Used by permission of Koch Engineering Co., Inc., Bull. KS-I.

Figure 14-54

Typical pressure dropmcapacity curves for countercurrent gas-liquid operation, for Koch Flexipac®.

Used by permission of Koch Engineering Co., Inc., Bull. KFP-4. Parameter lines on charts are gpm/ft^2.

Distill a diethanolamine (DEA) - triethanolamine (TEA) mixture to produce a 99.0 wt% DEA distillate product and a 95.0 wt% TEA bottoms product. The design material balance is as follows:

	Feed		Distillate		Bottoms	
	lb/mol	**mol/hr**	**lb/mol**	**mol/hr**	**lb/hr.**	**mol/hr**
DEA	5,000	47.6	4,740	45.1	260	2.5
TEA	5,000	33.5	50	0.3	4,950	33.2
Totals	10,000	81.1	4,790	45.4	5,210	35.7

To preclude thermal degradation of the amine solution, the reboiler pressure must be limited to 10 mm Hg absolute pressure. Total allowable pressure drop is 3 mm Hg.

Solution

Distillation calculations result in a reflux ratio L/D = 0.8, with 4 theoretical stages/trays for rectification and 4 theoretical trays for stripping, or a total of 8 trays. The design heat balance (neglecting heat losses) is as follows:

Heat Out	lb/hr	State	°F	Btu/lb	Btu/hr.
Distillate	4,790	Liquid	296	160	766,000
Bottoms	5,210	Liquid	381	240	1,250,000

Ovhd. Cond.	-	-	-	-	2,470,000
	10,000				4,486,000

Heat In					
Feed	10,000	Liquid	314	180	1,800,000
Reboiler (Q_r)		By Difference			2,686,000
	10,000				4,486,000

Column Sizing for Koch Sulzer Packing

Bottom Section

The bottom section vapor loading is controlled by the reboiler vapor where:

The design mass velocity at 10 mm Hg. (Figure 14-52) is 460 lb/hr-ft^2. Applying the "f" factor correction for six carbon atoms (Figure 14-53) results in

f = 1.12

A = 14,000 (1.12)/460 = 34.1 ft^2

D = 6.6 ft, use 6 ft. 9 in.

= 14,000/3,600 (35.8) (0.0031) = 35 ft/sec.

Considering the reboiler as a theoretical tray, for 3 theoretical trays the packing height for an HETP of 7 in. is 7 (3)/12 = 1.75 ft. This

requires 4 standard elements which is 6.7 in. deep, or 80 ft^3 of packing.

From Figure 14-51, the pressure drop is 0.38 mm Hg/ft or 0.85 mm Hg. total.

Top Section

The top vapor load at an L/D of 0.8 is 8,620 lb/hr. The total pressure drop for 7 theoretical trays is slightly less than 2 mm Hg and ρ_V is 0.002 lb/ft^3 at 8 mm Hg top pressure. Duplicating the calculations made for the bottom section results in:

D = 5 ft. 9 in.
Packing height = 2.8 ft.
Pressure drop = 1.06 mm Hg.
Therefore, use column diameter = 6 ft. 9 in.

Nomenclature

A = crosssectional area, ft^2
D = diameter, ft.

G = vapor rate, lb/sec. ft^2
g_c = 32.174 (lb_m/lb_f) (ft/sec^2)
L = Liquid rate, lb/sec. ft^2
L/D = reflux ratio.
P_f = packing factor
Q_c = condenser duty, Btu/hr.
Q_r = reboiler duty, Btu/hr.
V_r = reboiler vapor rate, lb/hr.
V_s = superficial vapor velocity, ft/sec.
P_1 = liquid density, lb/ft^3
ρ_V = vapor density, lb/ft^3

l_v = latent heat of vaporization, Btu/lb

Koch Flexipac® Structured Packing

This comes in four sizes, Types 1 through 4, and is constructed of corrugated metal sheets (see Figure 14-6GG). The types vary by corrugation size; the larger the type number, the greater is the depth of corrugation. The deeper corrugations give higher capacity and lower pressure drop. According to Koch reference [101], at the same efficiency, in countercurrent gas-liquid operation, this packing has a higher capacity and lower pressure drop than any available dumped or structured packing. The terminology for Figure 14-54 is

(14-84)

where

A = crosssectional area, ft^2
G = vapor rate, lb/hr.
V = vapor velocity, ft/sec.
ρ_V = vapor density, lb/ft^3
Δ_P = pressure drop, in. water/ft height.

Chart parameter lines are gpm/ft^2 crosssection.

The Flexipac® structural packings have better efficiency than available random packing, particularly at low liquid rate, see Reference [101].

Intalox High Performance Metal Structural Packing [102]

According to the manufacturer's literature [102], this packing surpasses the best other sheet-metal structured packings in terms of efficiency and capacity. See Figure 14-6II. The unique surface-

texturing feature provides greater use of the packing surface to achieve enhanced levels of mass transfer, and the overall geometry allows greater capacities and efficiencies to be obtained. Tests have been conducted on this and other packings at the University of Texas at Austin's “Separation Research Program, Center for Energy Studies” for distillation capacity and efficiency, and published in Reference 103.

For good and uniform performance of any structured packing, it is essential to have uniform, consistent vapor and liquid distribution; therefore, much care must be given to the design details. See earlier discussion in this chapter.

For specific final performance sizing of a distillation column using Norton's Intalox® structured packing, the designer is referred to the manufacturer's technical representatives, and should not assume that the preliminary results obtained from manufacturer's bulletin that are included here will necessarily serve as a final design. As a preliminary examination of a design problem (used by permission of Norton Chemical Process Products):

1. Calculate flow parameter X

(14-85)

where

A = area, ft^2 tower crosssectional area

L = Liquid mass rate, lb/sec.

G = vapor or gas mass rate, lb/sec.

ρ_G = gas density, lb/ft^3 at conditions

ρ_L = liquid density, lb/ft^3 at conditions

2. Read chart, Figure 14-55 and obtain C_o, ft/sec, at packing type shown.

Figure 14-55

Capacity correlation for two types of Intalox® structured packing. Data range: $5 \leq \sigma \leq 73$ and $0.07 \leq \mu \leq 1.1$.

Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2 (1994).

3. Calculate efficient capacity, C_{SC}:

(14-86)

σ = surface tension, dynes/cm.
μ = liquid viscosity, cP
4. Calculate

V = superficial gas velocity, ft/sec., or m/sec. depending on the units used

5. Read pressure drop coefficient, F.
Read, F, for the approximate calculated C_S, ft/sec from Table 14-40 at calculated liquid rate, lb/hr.ft^2. Then read Y from IMTP Packing Pressure Drop chart, Figure 14-21G and then read curves showing pressure drop, (may require interpolation).

Table 14-40 Norton Intalox® Structured Packing Pressure Drop Coefficient F* for Type 2T Packing

Used by permission of Norton Chemical Process Products Corp., Bull. I-S-I-R.

Air/Water	Liquid, kg/h · m^2	9,760	24,400	48,800	73,200	97,600	122,000
	Liquid, Ib/h · ft^2	2,000	5,000	10,000	15,000	20,000	25,000
	Intalox structured packing 2T						
	F when C_s in m/s	205	215	250	300	325	365
	F when C_s in ft/s	19	20	23	28	30	34
These packing factors can be used to predict the pressure drop of Intalox Structured Packing 2T when used with the Generalized Pressure Drop Correlation as illustrated in Norton's Intalox High-Performance Separation Systems; Figure 14-21G this text.							

6. Typical HETP data are shown in Figure 14-56AFigure 14-56B and Figure 14-56C.

Figure 14-56A Performance test results using Norton's Intalox® structured packing.

Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

Figure 14-56B

Performance test results using Norton's Intalox® structured packing.

Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

Figure 14-56C

Performance test results using Norton's Intalox® structured packing.

Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

Gempack® Structured Packing; Glitsch, Inc.

These packings are represented in part by Glitsch, Inc's bulletin [104] (Figure 14-6KK) and they cite the performance of the packing as:

1. High efficiency: 5–25 in. HETP (7.9 – 1.6 theoretical plates per meter).
2. Low pressure drop: 0.5–0.7 mm Hg/theoretical plate.
3. High vapor loadings: up to 0.5 C-Factor (up to 4.0 F-factor).
4. High liquid capacity: 0.5–55 U.S gpm/ft^2 (1.22–134 m^3/m^2/hr.).
5. Wide turndown ratio: limited only by distributors.
6. Availability: diameters from 4 in. to over 45 ft (10 cm to 13.7 meters).
7. Series available:
    a. Series AT: General purpose, suitable for low wetting rates and high vacuum applications.
    b. Series AS: General purpose, suitable for high wetting and heat transfer applications.
    c. Series AW: Latest generation of Gempack®, better efficiency with same hydraulic capacity as other series listed.
    d. Series AL: Especially suited for low wetting rates and

chemical applications.

e. Series BG: Wire gauze packing, crimped to 60° from horizontal, for high efficiency applications and very low wetting rates in clean service.

The performance and design information in Glitsch reference [104], as for all the other manufacturers with respect to either data and charts, is proprietary, but not necessarily warranted to be suitable for the designer's service/applications unless verified by the manufacturer's representatives.

Example 14-9: Heavy Gas-Oil Fractionation of a Crude Tower Using Glitsch's Gempak® (used by permission of Glitsch, Inc. Bulletin 5140)

The heavy naphtha-light gas oil fractionation zone of a crude tower has to be revamped to handle 25% more capacity. Because trays would be working at high percent flooding, Gempak structured packing is considered (Figure 14-56AFigure 14-56B and Figure 14-56C).

The loads are as follows:

Vapor rate	688, 000 lb/hr
Liquid rate	381,000 lb/hr
Vapor density, (ρ_V)	0.355 lb/ft^3
Specific gravity of gas oil	0.68
Liquid density, (ρ_L)	42.42 lb/ft^3
Liquid viscosity,	0.281 cP
NonFoaming System	
Tower diameter:	15 ft – 0 in.
Cross sectional area	176.72 ft^2
Height available for packed bed	6.5 ft (78 in.)

(excluding height for distributor)	

Solution

1. Calculate the vapor load:

Vapor, (688,000)/(3,600) (0.355) = 538.3 ft^3/sec.

Vapor velocity = V_s = (538.2)/(176.72) = 3.046 ft/sec.

2. Calculate the liquid load:

gpm = (381,000)/(8.33) (60) (0.68) = 1,120

gpm/ft^2 = (1,120)/(176.72) = 6.3

3. Select the Gempak size:

On the plot of C-factor vs. liquid load as shown in Figure 14-57 A mark the operating point (A) using 6.3 gpm/ft^2 and C-factor = 0.28. The point on the figure is above the Gempak AS (0.25-in. crimp height) packing flooding line. All other packing sizes would be operating below flooding. For example, if Gempak AS (0.375 in. crimp height) is to be used, the percent flooding at design loads will be:

- At constant L/V ratio – (OA/OD) × 100 = 75.2%
- At constant liquid rate – (100) (BA/BE) = 70.0%

Figure 14-57

For Example 14-8 typical Glitsch Gempak® preliminary design charts for structured packing. Note: These plots are to be used in carrying out the calculations described in the Glitsch bulletin. Graphic data presented here have been obtained in tests whose conditions may differ materially from your own.

Curves on these pages are for Gempak with crimp angles of 45° from the horizontal, and should be sufficient for a preliminary review. The actual surface texture, which determines the final design, should be chosen in consultation with the Glitsch process engineering staff.

Used by permission of Glitsch, Inc., Bul. 5140.

Normally in distillation, flooding at constant L/V ratio is more representative of actual plant operations (constant liquid loads may also be representative in cases like absorbers). In general, flooding percentages up to 75–80% are acceptable for continuous operation. Gempak AS (0.375-in. crimp height) would be a good selection for this example. Gempak AS (0.5-in. crimp height) would also be a good selection, with flooding of 70% at constant L/V ratio and 65% at constant liquid rate. Gempak AS (0.25-in. crimp height) would not be a good selection in this case since flooding would be over 100%.

4. Pressure Drop

On pressure drop plots, mark a C-factor of 0.28 and 6.3 gpm/ft^2 as shown in Figures 14-57B and 14-57C.
For Gempak AS (0.5-in. crimp height), ΔP = 0.31 in. liq/ft.
For Gempak AS (0.375-in. crimp height), ΔP = 0.55 in. liq/ft.
For a bed of 6.5 ft, the pressure drop for the bed is:

For Gempak AS (0.5-in. crimp height) = 0.31 × 6.5 = 2.0 inches of liquid.

For Gempak AS (0.375-in. crimp height) = 0.55 × 6.5 = 3.6 inches of liquid.

Pressure drop for both AS (0.5 in. crimp height) and AS 0.375-in. crimp height) packings is acceptable for crude tower operation.

5. Efficiency, Figure 14-57 D

On efficiency plot, mark C-factor at 0.28.
For Gempak AS (0.5-in. crimp height): HETP = 13.5 in. (0.89 NTS/ft)
For Gempak AS (0.375-in. crimp height): HETP = 10.2 in. (1.18 NTS/ft)
Note: The HETP's noted are valid only for o-/p xylene at the test conditions. Nevertheless, the ratio of HETP's should remain approximately constant.

Grid Packing: Nutter Engineering (Figure 14-6PP)

For applications and design details refer to the manufacturer concerning these types of packings. Figure 14-58 illustrates performance of No. 3 Snap Grid™. For mass transfer for distillation HETP, use:

(14-87)

where

(14-88)

HETP = height equivalent to a theoretical plate, inches
H_{og} = height of an overall gas phase transfer unit, inches
U_a = volumetric overall height transfer coefficient, Btu/(hr) (ft^3) (^oF)
m = slope of equilibrium line expressed in mole fraction.
G_m, L_m = gas, liquid molar rate based on superficial tower area, mol/(hr) (ft^2).
H_g, H_l = height of gas, liquid phase transfer unit, inches
Re = Reynolds number of gas phase

Figure 14-58

Nutter Snap-Grid™ typical performance charts for pressure drop.

Used by permission of Nutter Engineering, Harsco Corp., Bull. CSG-2, for Air-lsopar © 10gpm/ft^2.

Koch Flexigrid® Packing: Koch Engineering Co.

Koch Flexigrid® packing bulletin [106] states that the packing (Figure 14-6SS and Figure 14-6TT) has a fixed, ordered orientation and is supplied as layers that stack in a prescribed fashion within the bed. Features include [106]:

1. High capacity; constructed in 60 in. × 16 in. × 2¾ in. high modules.
2. High efficiency: constructed in 60 in. × 16 in. × 2¾ in. high modules.
3. Each successive layer of the grid is rotated 45° during installation.
4. Lower pressure drop.
5. Tendency to coke or foul far less than other grids, due to elimination of horizontal planes where liquids or solids can stagnate.
6. Low liquid holdup.
7. Fabricated of most metals, such as carbon steel, stainless steel, aluminum and others as required.

Capacity comparison at flooding is shown on Figure 14-59, as a function of the vapor and liquid capacity factors C_V and C_L. Koch developed correlations, used by permission [106]:

Figure 14-59

Comparison of capacities of Flexigrid® Styles 2 and 3 at flooding with 2-in. Flexiring® random packing, and a competitive grid.

Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

Flexigrid® Style 2 High Capacity

(14-89)

(14-89A)

Flexigrid® Style 3 High Efficiency

(14-90)

(14-90A)

const = constant.

A "viscosity correction" should be made if by multiplying the "% flood" obtained from Equations 14-89 through 14-90A by the term " " in cP.

(14-91)

(14-92)

where

A = tower area, ft^2

V_V = vapor rate, ft^3/sec.

V_L = liquid rate, U.S. gpm

ρ_L = liquid density, lb/ft^3
ρ_V = vapor density, lb/ft^3

Typical pressure drops are shown in Figure 14-60. Mass transfer and heat transfer evaluations should be referred to the manufacturer.

Figure 14-60

Pressure drop for Styles 2 and 3 Flexigrid® at selected liquid rates.

Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

Glitsch-Grid™ [107] (Figure 14-6UU)

This is an open area packing with multiple layers of lattice-type panels. This grid, as described by the manufacturer's bulletin, consists of vertical, slanted, and horizontal planes of metal. The vertical strips have horizontal flanges oriented alternately right and left. Due to the random overlap, the vapor path must zig-zag through the bed.

The manufacturer's data states that the grid has extremely low pressure drop (0.5 mm Hg/ft) at capacities higher than is possible with any other mass transfer device. Grid capacity is approximately 50% greater than conventional trays, and about 35% greater than 3½-in. ballast rings. The grid is highly resistant to fouling, plugging, or coking by tars or solids. See Figure 14-61A and Figure 14-61B for pressure drop and capacity performance comparison. HETP is available from the manufacturer and final design performance must be obtained from the same source.

Figure 14-61A Pressure drop comparison of Glitsch Grid™ and Ballast® rings at 10 gpm/ft^2.

Used by permission of Glitsch, Inc. Bull. 14-72.

Figure 14-61B

Capacity performance comparison for Glitsch-Grid™ and Ballast® rings.

Used by permission of Glitsch, Inc. Bull. 14-72.

14.30.1. Structured Packing: Technical Performance Features

Fair and Bravo [108] have performed extensive studies on structured packing, and have developed general models for flooding, pressure drop, and mass transfer. Structured packing is now generally considered to be cost effective for moderate pressure and vacuum distillation when compared to trays and random packings [108]. The test work of the authors considered the trade-named structured packings of Intalox®, Gempac®, Flexipac®, Mellapak®, Sulzer, and Montz in their studies. See earlier figures for installations of these packings, many of which are quite similar. All of the cited packings are corrugated sheet type designs, except the Sulzer, which is a fabricated wire gauze construction. Table 14-41 summarizes the characteristics of the selected packings. Refer to the respective manufacturers for confirming details and design application techniques.

Table 14-41 Characteristics of Representative Structured Tower Packings*

Used by permission of The American Institute of Chemical Engineers, Fair, J. R. and Bravo, J. L., *Chem. Eng. Prog.* Vol. 89, No. 1 (1990), p. 19; all rights reserved.

	Flexipac-2	**Gempak 2A**	**Intalox 2T**	**Montz B1-200**	**Mellapak 250Y**	** **Sulzer BX**
Specific area (m^{-1})	223	223	220	200	250	500
Void fraction	0.93	0.95	0.97	0.94	0.95	0.90
Corrugation angle (degrees)	45	45	45	45	45	60
Crimp height (m)	0.0125	0.0122	0.0104	0.0149	0.0119	0.0064
Corrugation side (m)	0.0177	0.0180	0.0223	0.0250	0.0171	0.0088
Corrugation base (m)	0.0250	0.0268	0.0390	0.0399	0.0241	0.0128

*All packing types listed are available in several different sizes. The corrugation angle is measured from the horizontal. Comfirm details with manufacturer.

**Wire gauze for comparison.

Flooding

At flooding or near flooding conditions [108]:

1. A rapidly increasing pressure drop with a relatively slight increase in gas rate (hydraulic flood) develops.
2. A rapidly decreasing efficiency with a relatively slight increase in gas rate (mass-transfer limitation) develops.
3. A general lack of column stability develops.

All of these conditions do not necessarily occur at the same liquid-gas loading. Generally, the mass-transfer limitation develops before the hydraulic flood condition, as loadings are increased. Fair and Bravo

[108] used the mass-transfer limitation as the limiting case for reasonable design of mass-transfer efficiencies. Figure 14-62 is based on hydraulic flood for several structured packings. The capacity limit is related to the corrugated elements as reflected in specific surface area. The capacity parameter, C_s in m/sec, = U_o.

(14-93)

V = superficial velocity, m/s.
ρ_G = gas density, kg/m^3
ρ_L = liquid density, kg/m^3

Figure 14-62

Flooding data for structured packings as reported by Billet [109]. Numbers following packing type indicate specific surface area in m^2/m^3.

Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R. and Bravo, J. L., *Chemical Engineering Progress*, V. 86, No. 1 (1990) p. 19; all rights reserved. Note, U_0 = vapor velocity, meters/sec.

Figure 14-63A and Figure 14-63B illustrate, for a specific, packing the hydraulic flood and mass-transfer efficiency limitations. The differences in crimp height can influence the results. Figure 14-63B shows the effect of a higher flow parameter taken using larger columns; the system apparently was approaching critical, but the cause of the performance drop is not yet known.

Figure 14-63A Flooding data for structured packings obtained by pressure drop measurements as well as by efficiency measurements (see Ref. 108 for sources).

Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R. and Bravo, J. L., *Chemical Engineering Progress*, V. 86, No. 1 (1990) p. 19; all rights reserved.

Figure 14-63B

Capacity correlation of structured Intalox® 2T. The capacity parameter is at the maximum efficient capacity (MEC) point defined by Rukovena and Koshy [151]. The MEC point is lower than the flood point and located where the packing will maintain its desired efficiency.

Used by permission of Rukovena, E. and Koshy, T. D., private communication, and *Ind. and Eng. Chem. Res.* V. 32, No. 10 (1993) p. 2400; all rights reserved.

Pressure Drop

Structured packings maintain mass-transfer performance with minimum pressure drop [108]. Two models are presented for calculating pressure drop: (1) Bravo-Rocha-Fair [111] and (2) Stichlmair-Bravo-Fair [112]. Each method is quite involved, with rather complex equations to calculate the ultimate pressure drop. The authors [108] recommend using the following in design:

1. Flooding: Empirical plots of data, as shown in Figure 14-62Figure 14-63AFigure 14-63B and Figure 14-64.

Figure 14-64

Pressure drop at the flood point as a function of loading. Values are calculated using the Stichlmair *et al.* model and distillation conditions using cyclohexane/n-heptane and Gempak® 2A packing. Note: ordinate $(N/m^2/m)/3.385^3$ = in. Hg/m.

Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R. and Bravo, J. L., *Chemical Engineering Progress*, V. 86, No. 1 (1990) p. 19; all rights reserved.

2. Pressure drop: Generalized model of Bravo *et al.* [111], in combination with Stichlman [112]. Pressure drop at the flood point as a function of the flow parameter at the flood point is given in Figure 14-62 [108] for several packings, using the Stichlman model equations for various system pressures [108].
3. Mass transfer: Generalized model of Bravo *et al.* [113], with discount factors for wetted surfaces.

Hatfield [114] describes the improvements in commercial performance when Pall rings were replaced with Goodloe® packing.

Pressure drop through gauze and sheet metal structured packings [115] applies for the region below the loading point, and cannot predict the flood point because liquid holdup vs. gas velocity is not included. The latest version of the equation is in Reference 108:

(14-94)

(14-95)

(14-96)

(14-97)

where

C_o = constant, value of 3.08 in. Equation 14-94 recommended for all structured packings similar to such types as Flexipac 2® and Gempak 2®. Note: Reference 108 provides values for other structured packing sizes and styles and wire gauze. Refer to manufacturer for confirmation.

where c, C_o, C_1, C_2, C_3 = correlation constants

Fr_1 = liquid Froude number

g = gravitational constant

Re_g = Reynolds number for gas

S = length of corrugation side

U_{ge} = effective velocity of gas

U_g = superficial velocity of gas

U_1 = superficial velocity of liquid

Δ_p = pressure drop per unit packed height

ε = packing void fraction

θ = angle of flow channel (from horizontal)

μ = viscosity

ρ = density

f (subscript) = flooding conditions

g (subscript) = gas

l (subscript) = liquid

Billet [123] reports that metal gauze-type packing (such as Goodloe®) gives better performance and lower costs for high vacuum distillations, particularly for thermally unstable mixtures. This comparison was made against Pall rings. For the performance information referenced here, see Figure 14-65 [123].

Figure 14-65

Performance of structured gauze packing vs. Pall rings.

Used by permission of Billet, R., *Chemical Engineering*, V. 79, No. 4 (1972) p. 68; all rights reserved.

For general references during vacuum operations:

1. Due to the rapid decrease in specific efficiency with increasing load, the optimum load factor is equal to a vapor capacity factor, that in vacuum systems is less than 60%–70% of the corresponding flood point (the actual decrease depending on the mixture). The average optimum load factor is given by [123]:

(14-98)

2. Specific efficiency of packing:

3. Average pressure drop per theoretical tray:

where

H = height of packing, m

n_t = number of theoretical trays

Δp = pressure drop, mm water column

u_V = vapor velocity per unit of free column crosssection, m/s.

ρ_v = density of vapor, kg/m^3

14.30.1. Guidelines for Structured Packings

1. Pressure drop: Range usually 0.3 – 0.5 in. water/ft, or 30-40 mm water/m of packed height [116].
2. Lower hold-up of liquid: Less than other packings; may require special attention to operational controls.
3. More difficult to purge: Operating vapors more difficult to purge from the system than random packings and trays. This poses potential fire, explosion, and toxic hazards.
4. Lower pressure drop for vacuum systems: Allows for better low pressure and vacuum system operation, and lower bottoms temperature, with less degradation of bottoms product.
5. Internal uniform distribution: Properly designed devices to distribute and redistribute liquid entering the column is critical to obtain best performance of these types of packings.
6. Materials of construction: The materials of fabrication for this type of packing are more critical to long life due to pad size, wall thickness of metal or plastic components, and actual selection based on the system corrosion, hydrogen attack, and oxygen attack in the column environment.
7. HETP values: Small packing with narrow corrugations gives low HETP values, but usually higher pressure drops. Plastic and some metal packing sheets, wires or corrugations may require special surface treatment to ensure good wettability.

8. Preliminary designs: These can usually be made from the generalized literature references; however, they are not a substitute for direct design of the specific system in question by the packing manufacturer, referencing to a specific packing size and style. Several competitive designs are often helpful for a final design selection.

The test studies of several different structured packings (Dean *et al.* [117]) indicated that structured packing performs well in water removed by triethylene glycol (TEG) from natural gas from field wells, giving a design point of $F_s = 3$ (where , V_s = superficial velocity, ρ_g = gas density) for sizing. A high efficiency drip-point distributor is recommended, because it was far superior to a notched trough or spray nozzle distributor. Excellent turndown to 12:1 was shown, still meeting outlet gas low vapor specifications. HETP values ranged from 44 in. to 102 in. in this application. The number of theoretical trays was 2.8 to 4.6, varying with liquid and gas rates and packing style/size. Flooding occurred at 2.28 to 3.98 F_s value at 0.7 gpm/ft^2 at 650 psig.

A commercial 4-ft diameter refinery depropanizer unit's performance after replacement of the lower half of a trayed tower with Intalox-2T® structured packing is described in Reference 118.

Nutter [141] Montz B-1 structured packing, licensed from Julius Montz GmbH of Hilden, Germany, Figure 14-6LL and Figure 14-6MM, has uniform sinusoidal corrugations, which avoids sharp corners. The embossing on both sides of the sheet metal is closely spaced projections in a dot-matrix pattern, which spreads the liquid in a uniform film [141], without holes or slots. The design allows for an

effective liquid-vapor seal at the tower wall by means of a metal wiper band. A representative performance is shown for pressure drop, Figure 14-66 and HETP, Figure 14-67. This packing is effectively used in low pressure (or vacuum) columns where large theoretical stages and low pressure drop are required, producing the lowest possible pressure drop per stage. Style B-1 is embossed sheet metal; Style BSH, is expanded textured sheet metal, high efficiency with maximum surface area and Style A-3 is wire gauze construction.

Figure 14-66

Representative pressure drop for Montz high-efficiency structured packings (several designs/styles) for a sheet metal packing. Performance can be accurately calculated by the manufacturer (See Figure 14-6LL and Figure 14-6MM). Used by permission of Nutter Engineering Co., Harsco Corp., Bull. B-I. Specifications and descriptions used were in effect when this publication was approved for printing. Nutter Engineering reserves the right to discontinue models or options at any time, or change specifications, equipment or designs without notice and without incurring obligation. Using any of this information for specific applications should be done in consultation with Nutter Engineering personnel.

Figure 14-67

Representative HETP for Montz high-efficiency structured packings (several designs/styles) for a sheet metal packing. Used by permission of Nutter Engineering Co., Harsco Corp., Bull. B-1. Specifications and descriptions used

were in effect when this publication was approved for printing. Nutter Engineering reserves the right to discontinue models or options at any time, or change specifications, equipment or designs without notice and without incurring obligation. Using any of this information for specific applications should be done in consultation with Nutter Engineering personnel.

14.30.2. Structured Packing Scale-up

Applications of structured packing in ethylene plant's various column systems [119] have been successfully achieved, but the individual manufacturers must be consulted to use their most directly applicable pilot and commercial data, which are generally not published. The use of published general correlations should only be used for a “first” or approximate design, while the delicate or important final design must be performed in cooperation with the manufacturer.

Hufton *et al.* [120] tested gauze type packing for large commercial units, which resulted in a scale-up procedure that is too involved to reproduce here.

1. Estimate required packing height from Bravo *et al.* [113] or manufacturer's information.
2. Calculate commercial scale HETP from Fair *et al.* [108] and/or Hufton *et al.* [120].

14.31. New Generalized Pressure Drop Correlation Charts

Many generalized pressure drop correlation charts (GPDC) have been developed over the years for random and structured packings, as shown in Figure 14-21CFigure 14-21DFigure 14-21EFigure 14-21FFigure 14-21GFigure 14-21H and Figure 14-21I. Recently, Kister *et al.* [174] developed GPDCs for new structured and random type packings. These charts are based upon the realization that liquid drainage in structured packings was restricted at the element to element transition rather than within each element. This implies that

the liquid accumulation which leads to flood is initiated at the element transition region. A fourth–generation of structured packings were developed, in which the main body of each element has layers inclined at 45°, but the top or both ends of each element are rounded or vertical to promote drainage at this point. These high-capacity structured packings were observed to offer more capacity than equivalent 45° inclined packings. Figure 14-68Figure 14-69Figure 14-70Figure 14-71Figure 14-72Figure 14-73Figure 14-74Figure 14-75Figure 14-76Figure 14-77Figure 14-78Figure 14-79 and Figure 14-80 show charts of the recent GPDC. See Kister *et al.* [174] for further details.

Figure 14-68

Strigle's GPDC chart for random packings.

Copyright® Gulf Publishing, Houston, TX (1994).

Figure 14-69

Kister and Gill's GPDC (SP) chart for structured packing. Kister et al., CEP, p. 28, (July 2007). By permission of AIChE, All rights reserved.

Figure 14-70

GPDC interpolation chart for Mellapak Plus 752Y with flood and pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-71

GPDC interpolation chart for Mellapak Plus 252Y with flood and pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-72

GPDC interpolation chart for Flexipac HC 2Y with pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-73

GPDC interpolation chart for Montz B1-250M with flood and pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-74

GPDC interpolation chart for Flexipac 3X with pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-75

GPDC interpolation chart for Flexipac 2Y with flood and pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-76

GPDC interpolation chart for # 1.5(M) Raschig Super-Ring with pressure drop data.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-77

GPDC interpolation chart for # 1.5(M) Raschig Super-Ring with a grossly excessive packing factor.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-78

Incorrect application of GPDC beyond the flood point.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-79

(a) Flexipac 1Y metal structured packing pressure drop vs. C-factor (air–water, 10 gpm/ft^2, Koch/Koch-Giltsch data). Kister et al., CEP, p. 28 (July 2007), by permission of AIChE, all rights reserved. (b) Flexipac 2Y metal structured packing pressure drop vs. C-factor (air–water, 10 gpm/ft^2, Koch/Koch-Giltsch data).

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

Figure 14-80

GPDC interpolation chart for packing in tower stripping section, showing an operating point above flood.

Kister et al., CEP, p. 28, (July 2007), by permission of AIChE, all rights reserved.

14.32. Mass and Heat Transfer in Packed Towers

Most packed towers are used for mass transfer operations such as absorption, distillation, and stripping; however, there are other uses, such as heat transfer quenching and entrainment knockout.

The standard packings and auxiliary features associated with these towers have been presented in connection with pressure drop considerations.

Because the packed tower is a continuous contacting device, in contrast to the stepwise plate tower, performance capacity is expressed as the number of transfer units, N, the height of the transfer unit, H.T.U., and mass transfer coefficients $K_G a$ and $K_L a$. Figure 14-81 identifies the key symbols and constant flow material balance.

Figure 14-81

Countercurrent packed tower symbols.

14.33. Number of Transfer Units, N_{OG}, N_{OL}

The transfer of mass between phases in a packed tower occurs either as essentially all from gas film controlling, all from liquid film controlling, or some combination of these mechanisms (see Figure 14-82). To express the ease (low number of transfer units) or difficulty of the transfer under the conditions of operation with respect to system equilibrium, the number of transfer units N_{OG} or N_{OL} required for the system is evaluated. These can be determined experimentally and the data used for similar systems. However, it is also important to be able to estimate the number of transfer units for systems where data are not available.

(14-99)

where N_{OG} = number of transfer units, based on overall gas film coefficients

N_{OL} = number of transfer units, based on overall liquid film coefficients

Z = height of packing, ft

H_{OG} = height of transfer units, based on overall gas film, coefficients, ft

H_{OL} = height of transfer unit, based on overall liquid film coefficients, ft

Figure 14-82

Mass transfer diagrams. The number of transfer units can be determined by the difference in concentration or vapor pressure, particularly over ranges where the equilibrium line is essentially straight. **Used by permission of Czermann, J. J., Gyokheqyi, S. L., and Hay, J. J., *Petroleum Refiner*, V. 37, No. 4 (1958) p. 165; all rights reserved.**

The transfer process is termed gas film controlling if essentially all of the resistance to mass transfer is in the gas film. This means that the gas is usually quite soluble in, or reactive with, the liquid of the system. If the system is liquid film controlling, the gas is relatively insoluble in the liquid and the resistance to transfer is in the liquid film. Many systems are a combination of the two. Without good data it is next to impossible design of equipment exactly, although satisfactory designs are possible. System information is presented in Table 14-42 as a guide. Other data for different systems exist in the literature in a scattered fashion.

Table 14-42 System Film Control*.

Gas film
*From: M. Leva, *Tower Packings and Packed Tower Design*, 2nd Ed. p. 91, U.S. Stoneware Co. (1953), by permission, now, Norton Chemical Process Products Corp.

1. Absorption of ammonia in water

2. Absorption of ammonia in aqueous ammonia

3. Stripping of ammonia from aqueous ammonia

4. Absorption of water vapor in strong acids

5. Absorption of sulfur trioxide in strong sulfuric acid

6. Absorption of hydrogen chloride in water

7. Absorption of hydrogen chloride in weak hydrochloric acid
8. Absorption of 5 vol. percent ammonia in acids

9. Absorption of sulfur dioxide in alkali solutions

10. Absorption of sulfur dioxide in ammonia solutions

11. Absorption of hydrogen sulfide in weak caustic

12. Evaporation of liquids

13. Condensation of liquids

Liquid film

1. Absorption of carbon dioxide in water

2. Absorption of oxygen in water

3. Absorption of hydrogen in water

4. Absorption of carbon dioxide in weak alkali

5. Absorption of chlorine in water
Both gas and liquid film

1. Absorption of sulfur dioxide in water

2. Absorption of acetone in water

3. Absorption of nitrogen oxide in strong sulfuric acid

For (1) dilute solutions or (2) equal molar diffusion between phases (e.g., distillation)

(14-100)

where $(y - y*)$ = driving force, expressed as mol fractions

y = mol fraction of one component (solute) at any point in the gas phases of the contacting system

$y*$ = mol fraction gas phase composition in equilibrium with a liquid composition, x

x = mol fraction in the liquid at the same corresponding point in the system as y

1, 2 = inlet and outlet conditions of system

If the system has more than two components, the calculations may be based on the component which varies the most in passing through the unit, or the component for which good data are available.

A large majority of systems have operating lines and equilibrium curves which can be assumed to be straight over the range covered by the design. For the conditions of a straight line equilibrium curve, $y* = mx$, Colburn [10,11] has integrated the relation above to obtain:

(14-101)

where N may be N_{OG} or N_{OL} depending on the operation.

Table 14-43 identifies several important conditions that affect the values of P″ and M. These are extracted from Colburn's larger summary [11].

Table 14-43 Values to use with Transfer Equation and Figure 14-68

By permission, A P. Colburn, *Ind. Eng. Chem.* 33, 459 (1941). The American Chem. Soc., all rights reserved.

Condition of operation	P"	M
Absorption:		
1. Constant mG_m/L_m	mG_m/L_m	
2. Varying mG_m/L_m	m_2G_{m2}/L_{m2}	
Desorption (stripping):		
3. Constant L_m/mG_m	L_m/mG_m	
4. Varying L_m/mG_m	L_{m2}/m_2G_{m2}	
Distillation, enriching[1]		
5. Constant mG_m/L_m		Same as 1
6. Varying mG_m/L_m stripping, closed steam[2]		Same as 2
7. Constant L_m/mG_m	L_m/mG_m	
8. Varying $L_m,/mG_m$ stripping, open steam[2]	L_{m2}/m_2G_{m2}	
9. Constant L_m/mG_m	L_m/mG_m	x_1/x_2
10. Varying L_m/mG_m	L_{m2}/m_2G_{m2}	

*Equilibrium value
Subscripts 1 and 2 refer to the concentrated and dilute ends of the unit respectively
[1]Concentrations and m are based on high boiler or “heavy key”
[2]Concentrations and m are based on low boiler or “light key”
m = slope of equilibrium line (mol-fraction solute in gas)/(mol-fraction solute in liquid)

Figure 14-83 is a plot to aid in solving the equation for N (or N_{OG} or N_{OL}).

Figure 14-83

Colburn Plot for transfer units.

Reproduced by permission of the American Chemical Society; Colburn, A. P., *Ind. Eng. Chem.* V. 33 (1941) p. 459; all rights reserved.

For constant temperature absorption, with no solute in the inlet liquid, $x_2 = 0$, and the abscissa becomes y_1/y_2.

For Concentrated Solutions and More General Application

The following equation applies for diffusion in one direction (e.g., absorption, extraction, desorption) [74]:

(14-102)

(14-103)

or

(14-104)

or

(14-105)

where $(1 - y)_m$ = log mean average of concentration at the opposite ends of the diffusion process, $(1 - y)$ in main gas body, and $(1 - y*)$ at the interface [74]

y = Concentration of solute in gas, mol fraction

$y*$ = Concentration of solute in gas in equilibrium with liquid, mol fraction

Y = Concentration of solute in gas, lb mol solute/lb mol solvent gas

$Y*$ = Concentration of solute in gas in equilibrium with liquid, lb mol solute/lb mol solvent gas

If the liquid film controls:

(14-106)

(14-107)

where x = concentration of solute in liquid, mol fraction

x^* = concentration of solute in liquid in equilibrium with gas, mol fraction
X = concentration of solute in liquid, lb mol solute/lb mol solvent
X^* = concentration of solute in liquid in equilibrium with the gas, lb mol solute/lb mol solvent

It is usually necessary to graphically integrate the first terms of the above equations, although some problems do allow for mathematical treatment.

Example 14-10

A gas stream comprising 92 mole % dry air and 8 mole % SO_2 is to be scrubbed with fresh water in a packed tower. The feed flow rate will be 60 kmol/h of gas and the exit gas should contain less than 0.18% SO_2.

The tower is to be designed to operate at 70% of flooding velocity. Operation will be isothermal at 30°C and 1 bar. The water flow rate is to be twice the minimum flow rate.

Calculate:

The tower diameter.

The packed height required, assuming the height of an overall gas phase transfer unit is 0.9m.

Data:

Equilibrium relationship:	$y* = 25.4x$
Packing factor	$a_p = 253\ m^2/m^3$
Voidage	$e = 0.79$

Density of water	
Acceleration due to gravity	$g = 9.81\ m/s^2$
Molar mass of air	29.0 kg/kmol
Molar mass of SO_2	64 kg/kmol
1 kmol of gas at 0°C and 1 bar occupies 22.4 m^3	

Solution

$L = 2\ L_{min}$

$u = 70\%\ u_{Flood}$

$y* = 25.4\ x$

$e = 0.79$

$a_p = 253$

= 55.2 kmol/h

For minimum flow:

Operating Line Equation:

Total Flow rate:

	= 1908 kg/h
	= 1600.8 kg/h
	= 52704 kg/h
	= 53011.2 kg/h

Densities:

Average molecular weight, Mw_{avg} = 0.08 x 64 + 0.92 × 29 = 31.8 kg/kmol

Density of water,

Density of gas, ρ_G = 1 kmol at NTP occupies 22.4 m^3

31.8 kg at 25°C occupies

Therefore, density of gas is:

From Figure 14-84, the intercept = 0.019

Figure 14-84

Generalized correlation for flooding rates in packed towers for Examples 14-10.

where $\mu_L = \mu_W$

The gas velocity at 70% of flooding velocity is: u = 0.7 × 0.5288 = 0.3699 m/s
Volumetric gas flow = 60 × 24.45 (kmol/h × m^3/kmol)
= 146.70 m^3/h
= 0.4075 m^3/s
Area of the tower = 0.4075/0.3699 = 1.1016 m^2

Tower diameter from the area:

(ii) The number of transfer unit is:

Packed height,
$z = 0.9 \times 6.07 = 5.46$ m.

Example 14-11: Number of Transfer Units for Dilute Solutions

An existing 10-in. I.D. packed tower using 1-inch Berl saddles is to absorb a vent gas in water at 85°F. Laboratory data show the Henry's Law expression for solubility to be $y* = 1.5x$, where $y*$ is the equilibrium mol fraction of the gas over water at compositions of x mol fraction of gas dissolved in the liquid phase. Past experience indicates that the H_{OG} for air-water system will be acceptable. The conditions are: (refer to Figure 14-81).

$G'_1 = 200$ mol gas/hr (ft^2)
$L'_2 = 500$ mol water/hr (ft^2)
$y_1 = 0.03$ (inlet)
$y_2 = 0.001$ (outlet)
$x_2 = 0$ (inlet)
$x_1 = ?$ (outlet)

Determine the number of transfer units, and the packed tower height.

Solution

Material Balance

Dilute solutions, assume constant L' and G'.

Gas phase change = $G_1 = (y_1 - y_2) = 200\ (y_1 - 0.001)$

Liquid phase change = $L_2 (x_2 - x_1) = 500 (0 - x_1)$

Because the (−) sign has no significance, except to indicate the direction of mass change, use 500 (x_1).

Now, to use the simplified

(14-100)

Assume values of y_1 and solve the equated mass change for values of x.

$200 (y_1 - 0.001) = 500 (x_1)$

Assume y	**Calculated x_1**	**Equilibrium y ∗**	
0.03	0.0116	0.0174	79.4
0.025	0.0096	0.0144	94.4
0.02	0.0076	0.0114	116.4
0.015	0.0056	0.0084	151.4
0.01	0.0036	0.0054	217.
0.005	0.0016	0.0024	384.
0.001	0	0	1000.

Example: Calculate x, as illustrated for point y = 0.015

Calculate the equilibrium values, $y*$, as illustrated for point of x_1 = 0.0116

$y* = 1.5\ (0.0116) = 0.0174$

Calculate $1/(y - y*)$ for point corresponding to: $x_1 = 0.0116$ where $y* = 0.0174$

Area Under Curve

Figure 14-85 is a summation of steps indicated, or the area can be circumscribed with a planimeter and evaluated.

Area = 6.27 units, then number of transfer units,

$N_{OG} = 6.27$

Figure 14-85

Graphical integration for Example No. 14-11.

Height of Transfer Unit

From data of Mehta and Parekh for 1-in. Berl Saddles [40],

$H_{OG} = 0.86$ ft

Height of packed section:

$Z = (6.27)\ (0.86) = 5.4$ ft

Total Packed Height Recommended

Use: 8.5 to 10 ft of 1-in. Berl saddles

To complete the design, the tower should be checked for loading and flooding conditions, and the pressure drop established. However, this procedure will not be repeated as it can be found elsewhere in this text.

Example 14-12: Use of Colburn's Chart for Transfer Units—Straight Line Equilibrium Curve, Figure 14-83 Constant Temperature Operation

Use the previous example on dilute solutions and solve by Colburn's Chart [11].

Abscissa: $y_1/y_2 = 0.03/0.001 = 30$, because $x_2 = 0$ (no dissolved solute gas in inlet water)

Parameter: $mG'/L' = (1.5)\ (200/500) = 0.6$

Note that m is the slope of the straight line equilibrium curve, m = 1.5

Reading chart: At $y_1/y_2 = 30$ and $mG'/L' = 0.6$

No. transfer units, N = 6.4

This compares with the value from graphical integration of 6.27 and is a good check.

If N = 6.4 were used for the tower:

= 6.4 (0.86) = 5.5 ft minimum for process operations

Example 14-13: Number Transfer Units—Concentrated Solutions

Using the basic problem for dilute solutions, assume the following conditions for a higher concentration.

$G' = 200$ mol gas/hr (ft^2)

$L' = 500$ mol water/hr (ft^2)

$Y_1 = 0.30$ (inlet)

$y_2 = 0.01$ (outlet)

$x_2 = 0$

$y* = 1.54$

Material Balance

Based on inert gas

Gas phase change =

Liquid phase change =

These changes must be equal:

Assume values of y and solve for corresponding values of x.

Let $y = 0.3$

Calculate equilibrium $y*$ from values of x.

Assume that for this concentration range (this will usually not be same as for dilute solution tower):

$y* = 1.5x$

$y* = 1.5\ (.1048) = 0.157$

Calculate $(1 - y)$:

$1 - y = 1 - 0.30 = 0.70$

Calculate $(1 - y*)$:

$1 - y* = 1 - 0.157 = 0.843$

Calculate $(1 - y)_m$, arithmetic average of non-diffusing gas concentration at ends of diffusing path:

Calculate $1/(y - y*)$:

Calculate $(1 - y)_M/(1 - y)(y - y*)$:

$= 0.7715(7.0)/0.7 = 7.715$

A plot of y versus $(1 - y)_m/(1 - y)(y - y*)$ gives the number of transfer units as presented in the previous example. As a second solution:

From Figure 14-86 the area under the curve for y versus $1/(y - y*)$ is 5.72 units = N_{qg} (approximate)

Figure 14-86

Graphical integration number of transfer units for Example 14-11.

The correction is:

Assumed	**Calc**					**1**	**$(1 - y)_M$**
y	**x**	**y∗**	**1 − y**	**1 − y∗**	**$(1 - y)_M$**	**y − y∗**	**(1 − y) (y − y∗)**
0.30	0.1048	0.157	0.70	0.843	0.7715	7.0	7.715
0.20	0.0631	0.0945	0.80	0.9055	0.8527	9.48	10.10
0.15	0.0446	0.0667	0.85	0.9333	0.8916	12.0	12.58
0.10	0.0276	0.0413	0.90	0.9587	0.9293	17.02	17.55
0.05	0.0142	0.0212	0.95	0.9787	0.9642	24.0	25.40

0.05	0.0142	0.0213	0.95	0.9787	0.9643	34.9	35.40
0.01	…	…	0.99	1.000	0.9950	100.0	100.50

Therefore:

$N_{OG} = 5.72 + 0.172 = 5.89$ transfer units

Note that graphical integration is never exact and hence the correction often makes little difference, except for cases of curved equilibrium lines.

From Figure 14-86 the area under the curve, y versus $(1 - y)_m/(l - y)(y - y*)$ is only slightly larger than the y versus $1/y - y*$ for this case. To avoid confusion the figure was only integrated for the latter. However, it could be performed for the former and the result should be very close to 5.89.

14.34. Gas and Liquid-phase Coefficients, k_G and k_L

Recent studies indicate that individual film transfer coefficients may be correlated with good agreement for Raschig rings and Berl Saddles for aqueous and nonaqueous systems [67]:

(14-108)

This has been shown to correlate for a wide variety of tower packings, various operating conditions, and physical properties of solute and inert gases. The calculated k_G must be used in conjunction with the effective interfacial areas determined by Shulman [65] (Figure 14-47),

to establish a reliable value for kGa. Figure 14-47 should be used with the abscissa as for inert gas other than air [67]:

(14-109)

where D_v = diffusivity of solute in gas, ft^2/hr
k_G = gas-phase mass transfer coefficient lb mol/hr (ft^2) (atm)
M_m = mean molecular weight of gas, lb/lb mol
P_{bm} = mean partial pressure of inert gas in the gas phase, atm
μ_{Ga} = gas viscosity, lb/(hr) (ft)
ρ_g = gas density, lb/ft^3
G' = superficial gas rate, lb/(hr)/(ft^2)

(14-110)

(14-111)

H' = Henry's law constant, lb mols/(ft^3) (atm)
k_L = liquid-phase mass transfer coefficient, lb mols/(hr) (ft^2) (lb mols/ft^3)

The relation

(14-112)

is reported to correlate within ±8% based on test data, and appears to be founded on a sound investigative program.

For the liquid phase based on Raschig ring and Berl saddle data [65]:

(14-113)

Use of k_G and k_L

1. From physical properties of system, determine k_G and k_L. If system is known, or can be assumed to be essentially all gas or all liquid film controlling, then only the controlling k need be calculated. For greater accuracy, both values are recommended, because very few systems are more than 80% controlled by only one k.
2. Combine effective interfacial area to calculate k_Ga or k_La.
3. Determine K_Ga by:

(14-114)

14.35. Height of a Transfer Unit, Hqg, H_{OL}, HTU

The earlier concept of height equivalent to a theoretical plate (HETP), which relates the height of packing to a unit of transfer known as the theoretical stage or plate, has generally been dropped in favor of the "height of a transfer unit" HTU. This is designated as H_{og} or H_{ol} depending on whether it was determined from gas or liquid film data. HETP data for absorption and distillation is given in the section under packed tower distillation.

Height of Overall Transfer Unit

For small changes in concentration and if the total number mols of gas and liquid remain essentially constant; the following is applicable to all but very concentrated solutions. For the latter case see References 18 and 74.

(14-115A)

(14-115B)

(14-116)

(14-117)

Height of Individual Transfer Unit

These calculations apply to the same limiting conditions as the preceding section. Some data are reported as individual gas or liquid film coefficients or transfer unit heights, however, it is often possible to use it as overall data if the conditions are understood.

(14-118)

(14-119)

(14-119)

(14-120)

(14-121)

where $A' = L_m/mG_m$

For predominately liquid film controlling systems, $A'H_G$ is almost negligible and $H_{OL} = H_L$; likewise for gas film controlling, H_L/A' is negligible and $H_{OG} = H_G$.

where G_m = gas mass velocity, lb mol/(hr) (ft^2)

L_m = liquid mass velocity, lb mol/(hr) (ft^2)

K_Ga = overall gas mass-transfer coefficient, lb mol/(hr) (ft^3) (atm)

K_La = overall liquid mass-transfer coefficient, lb mol/(hr) (ft^3) (lb mol/ft^3)

k_Ga = individual gas mass-transfer coefficient, lb mol/(hr) (ft^3) (atm)

k_La = individual liquid mass-transfer coefficient, lb mol/(hr) (ft^3) (lbmol/ft^3)

P_{ave} = average total pressure in tower, atmospheres

H_L = height of liquid film transfer unit, ft

H_G = height of gas film transfer unit, ft

a = effective interfacial area for contacting gas and liquid phases, ft^2/ft^3. Because this is very difficult to evaluate, it is usually retained as a part of the coefficient such as K_Ga, K_La, k_Ga, and

$k_L a$.
V′ = vapor flow rate, lb mol/hr

Estimation of Height of Liquid Film Transfer Units

The following relation is used in estimating liquid film transfer units [62]. For the proper systems H_L may be assumed to be equal to H_{OL}.

(14-122)

Where $\mu_{La}/\rho_L D_L$ = Schmidt number
Hl = height of transfer unit, ft
L’ = liquid rate, lb/(hr) (ft^2)
μ_L = viscosity of liquid, lb/(ft) (hr)
D_L = liquid diffusivity, ft^2/hr
ϕ and j are constants given in Table 14-44.

Table 14-44 Liquid Film Height of Transfer Unit*.

Packing	ϕ	j	Range of L′Lb/hr (ft^2)
Raschig Rings (in.)			
⅜	0.00182	0.46	400–15,000
½	0.00357	0.35	400–15,000
1	0.0100	0.22	400–15,000
1.5	0.0111	0.22	400–15,000
2	0.0125	0.22	400–15,000
Berl Saddles (in.)			
½	0.00666	0.28	400–15,000
1	0.00588	0.28	400–15,000
1.5	0.00625	0.28	400–15,000
3 In. Partition rings,			

stacked staggered	0.0625	0.09	3,000–14,000
Spiral Rings, stacked staggered			
3-in. single spiral	0.00909	0.28	400–15,000
3-in. triple spiral	0.0116	0.28	3,000–14,000
Drip-point grids (continuous flue)			
Style 6146	0.0154	0.23	3,500–30,000
Style 6295	0.00725	0.31	2,500–22,000
*Reproduced by permission, Treybal, R. E., *Mass Transfer Operations*, McGraw-Hill Book Co., Inc. (1955), p. 237, using data of Sherwood, T. K. and Holloway, F. A. L. [62] and of Molstad, McKinney and Abbey [51], all rights reserved.			

Diffusivity values are given in Table 14-45.

Table 14-45 Diffusion Coefficients of Gases and Liquids in Liquids at 68°F (Dilute Concentrations).
From: Perry J. H., *Chem. Engrs. Handbk.* 3rd Ed., p. 540, McGraw-Hill Book Co., Inc. © (1950). By permission.

	Solvent	**Diffusion coefficient, D_L ft^2/hr (multiply all values by 10^{-5})**
Gas		
Oxygen	Water	7.0
Carbon Dioxide	Water	5.82
Nitrous Oxide	Water	5.86
Ammonia	Water	6.8S
Chlorine	Water	4.74
Bromine	Water	4.66
Hydrogen	Water	19.92
Nitrogen	Water	6.37
Hydrogen Chloride	Water	10.25
Hydrogen		

Hydrogen Sulfide	Water	5.47
Acetylene	Water	6.06
Liquid		
Sulfuric Acid	Water	6.72
Nitric Acid	Water	10.15
Methanol	Water	4.97
Sodium Chloride	Water	5.23
Sodium Hydroxide	Water	5.86
Note: Additional data are given in the references, as well as the International Critical Tables.		

Estimation of Height of Gas Film Transfer Units

The relation [61, 62, 63]

(14-123)

describes a reasonable part of the gas film data. It allows the conversion of the ammonia-air-water data of Fellinger [27] to other systems. Table 14-46 gives the constants for the equation.

Table 14-46 Gas Film Height of Transfer Units [27, 50, 51]

From the data of Fellinger [27] and of Molstad et. al [50, 51] as presented in Treybel, R. E. *Mass Transfer Operations*, p. 239, McGraw-Hill Book Co. Inc. © (1955), Ref. 74, by permission, all rights reserved.

Packing	α	β	γ	G′	L′
Raschig Rings					
(in.)					
⅜	2.32	0.45	0.47	200–500	500–1,500

1	7.00	0.39	0.58	200–800	400–500
	6.41	0.32	0.51	200–600	500–4,500
1.5	17.3	0.38	0.66	200–700	500–1,500
	2.58	0.38	0.40	200–700	1,500–4,500
2	3.82	0.41	0.45	200–800	500–4,500
Bed Saddles					
(in.)					
½	32.4	0.30	0.74	200–700	500–1,500
	0.811	0.30	0.24	200–700	1,500–4,500
1	1.97	0.36	0.40	200–800	400–4,500
1.5	5.05	0.32	0.45	200–1,000	400–4,500
3-in. Partition Rings (Stacked (Staggered)	650	0.58	1.06	150–900	3,000–10,000
Spiral Rings					
(stacked staggered)					
3-in. single spiral	2.38	0.35	0.29	130–700	3,000–10,000
3-in. triple spiral	15.6	0.38	0.60	200–1,000	500–3,000
Drip Point Grids					
(continuous flue)					
Style 6146	3.91	0.37	0.39	130–1,000	3,000–6,500
Style 6295	4.65	0.17	0.27	100–1,000	2,000–11,500

α, β, γ = constants peculiar to packing for dilute and moderate concentrations [74]:

(14-120)

(14-121)

Figure 14-87 presents some of the data of Fellinger [168], as presented in Reference 40 for H_{OG} for the ammonia–air–water systems. This data may be used with the Sherwood relations to estimate H_L and H_G values for other systems.

Figure 14-87

Fellinger's overall gas film mass transfer data for ammonia-water system.

Used by permission of Leva, M. *Tower Packings and Packed Tower Design*, 2nd Ed. U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.) (1953).

Estimation of Diffusion Coefficients of Gases

Good, reliable diffusion data is difficult to obtain, particularly over a wide range of temperatures. The Gilliland relation is [63]:

(14-124)

where T = absolute temperature, °R
Ma, Mb = molecular weights of the two gases, A and B
P = total pressure, atm
V_A, Vb = molecular volumes of gases, obtained by Kopp's law of additive volumes, cm^3/gm mol at normal boiling point. See Table 14-47 or Volume 1.

Table 14-47 Atomic and Molecular Volumes.
By permission, R. E. Treybal, *Mass-Transfer Operations*, p. 27, McGraw-Hill Book Co., Inc. (1955).
Also see Ref. 63 for additional data.

Atomic volume		**Molecular**	**volume**
Carbon	14.8	H_2	14.3
Hydrogen	3.7	O_2	25.6
Chlorine	24.6	N_2	31.2
Bromine	27.0	Air	29.9
Iodine	37.0	CO	30.7
Sulfur	25.6	CO_2	34.0
Nitrogen	15.6	SO_2	44.8
Nitrogen in primary amines	10.5	NO	23.6
Nitrogen in secondary amines	12.0	N_2O	36.4
Oxygen	7.4	NH_3	25.8
Oxygen in methyl esters	9.1	H_2O	18.9
Oxygen in higher esters	11.0	H_2S	32.9
Oxygen in acids	12.0	COS	51.5
Oxygen in methyl ethers	9.0	Cl_2	48.4
Oxygen in higher ethers	11.0	Br_2	53.2
Benzene ring: subtract	15	I_2	71.5
Naphthalene ring: subtract	30		

Diffusion coefficients are used to estimate K_Ga values for gas film controlling systems:

(14-125)

Ammonia–air–water system data, Figure 14-87, is often used by converting H_{OG} (ammonia–air) to its corresponding K_Ga, and then substituting the above relation for the unknown K_Ga.

Example 14-14: Design of Ammonia Absorption Tower, Figure 14-88A and Figure 14-88B

An inert gas process vent stream contains 91 lb/hr (5.35 mol/hr) ammonia. This is 5.7% (volume) of the total. The absorber is to operate at 150 psig and recover 99% of the inlet ammonia as aqua using 90°F water. Average vapor mixture molecular weight = 11.6. Determine (a) the number of transfer units for the absorption, (b) height of the transfer unit using 1-in. Berl saddles, (c) the tower diameter, and (d) the water rate.

Figure 14-88A Tower specifications for Example 14-12

Figure 14-88B

Tower internal specifications, Spray or Packed Columns.

Solution

Material Balance

Assume production of 8 wt % aqua. Then:

Entering water, X_2, mol NH_3/mol H_2O = 0

Leaving water as aqua, X_1, mol NH_3/mol H_2O

Entering gas, Y_1, mol NH_3/mol inert gas

Leaving gas, Y_2, mol NH_3/mol inert gas

= (0.01) (.0607) = 0.000607

To calculate the equilibrium curve at 90°F (constant temperature) for the system aqua ammonia-ammonia-inert vapors follow the steps listed:

Heat of solution (-Q) = 45.8676 + n (286.103)-a'-nb', in kilojoules/g mole

∗n, g. moles H_2O	**∗-Q p.c.u./lb mole NH_3**	**-Q Btu/lb NH_3**	**x, mole fraction NH_3**	**Lb. NH_3 per 100# H_2O**	**∗Adiabatic Temp. rise, °F**
1	6,600	698	0.5	94.5	338
2.33	7,820	829	0.3	40.5	238
4	8,040	851	0.2	23.6	164
9	8,220	873	0.1	10.5	83
19	8,290	879	0.05	4.97	41.6
49	8,580	906	0.02	1.93	17.1

Data on heats of solution of ammonia taken from International Critical Tables Vol. V, pg. 213 [35] by Sherwood and Pigford *(Absorption and Extraction*, pg. 161, 2nd Ed., McGraw Hill Book Co., Inc.) Ref. 63.

1. Assume values for Y, (0.012 for example) mol ammonia/mol of inert gas and read corresponding vapor pressure of ammonia from Figure 14-89 (curve Y) (=102 mm). This figure was calculated from

Figure 14-89

Ammonia vapor pressure-inerts data at a fixed pressure.

for a total pressure system at 164.7 psia (150 psig) and a temperature of 90°F with vapor pressures read from published data, Figure 14-90.

Figure 14-90

Partial pressure of ammonia over aqueous solutions of ammonia.

Used by permission of Leva, M., *Tower Packings and Packed Tower Design*, 2nd Ed., U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.) (1953).

2. At the values of vapor pressure at 90°F (32.2°C) read the corresponding weight percent aqua ammonia (=8%).
3. Convert this weight percent ammonia to lb mols ammonia/lb mol water by

4. Plot equilibrium curve (curve A) of Figure 14-91. (at Y = 0.012, x = 0.092, example)

Figure 14-91

Equilibrium curves for ammonia-water; an operating system for production of 8% aqua (by weight) at total pressure of 150 psig.

If the temperature rise over the temperature range is very high, then operation at constant temperature requires internal cooling coils in the column, or other means of heat removal. Usually this condition will require considerably fewer transfer units for the same conditions, when compared to adiabatic operation.

To calculate the equilibrium curve whilst taking the heat of solution into account, i.e., operate adiabatically with liquid temperature variable, follow the steps:

1. Assume a temperature rise (for example, 17.8°F) and read from Figure 14-92 (temp. rise) the lb NH_3/100 lb H_2O (= 2, example).

Figure 14-92

Temperature rise due to heat of solution, ammonia in water.

2. Convert this lb NH_3/100 lb H_2O to lb mol NH_3/lb mol H_2O by

3. Convert this (0.0215) lb mol NH_3/lb mol H_2O to weight percent NH_3 by step (3) of previous paragraph (= 1.95%).

4. Read aqua-ammonia vapor pressure curves at wt percent NH_3 and corrected temperature (base temperature plus rise) (= 90 + 17.8 = 107.8°F, e.g., or 41.1°C). Read 29.5 mm Hg as vapor pressure of ammonia, Figure 14-90.

5. Read "Y" curve, Figure 14-89 at value of vapor pressure to get Y (= 0.0035 lb mol NH_3/lb mol inerts).

6. Plox X (= 0.0215) and Y (= 0.0035) to get equilibrium curve, which accounts for this effect of heat of solution Curve B, (Figure 14-91).

Determine Number of Transfer Units, N

The number of transfer units is determined graphically by:

1. Bisecting the vertical distance (line C) of Figure 14-91 between the 8% aqua operating line and the equilibrium curve B.
2. Starting at Y_2, draw horizontal line to line C. Extend this horizontal to right of C far enough to make "Y_2 to line C" equal to "line C to end," then step vertically to operating line, D. Move horizontally to line C, so that (a) (b) = (b) (c). Continue moving up the tower in this stepwise fashion.
3. At the end, X_1Y_1, if steps do not end at an exact point, estimate the fraction of vertical step required and report as fractional transfer unit.

This problem steps off six full transfer units and approximately ⅓ of the seventh. Report as 6.33 units.

Performance Interpretation

The point X_1 represents 0.092 mol NH_3/mol H_2O, which is equivalent to 8 wt% ammonia solution. If a weaker product were desired fewer transfer units would be required.

Point (e) is the intersection of the equilibrium curve and the operating line and represents the equilibrium condition for tower outlet liquor, and the maximum liquor concentration.

Note that the 10% (wt) aqua product operating line is shown on the diagram, but such a concentration cannot be reached when operating at 150 psig. A greater pressure is required in order to lower the equilibrium curve.

Curve A represents the equilibrium condition for water entering at 90°F, the gas entering saturated with water vapor at 90°F and isothermal tower operation.

Height of Transfer Unit

From Figure 14-87, the experimental H_{og} may be picked based on ammonia from a mixture with air absorbing in water. Assume an 18 O.D. tower (pipe) which has an I.D. of 16.8 in.

At inlet,

At outlet,

H_{og} = 1.6 ft (interpolated) based on ammonia–air–water system. The system under study has inerts other than air.

Tower Height Based on Air as Inert Gas in System

$$Z = NH_{OG}$$

$$Z = (6.33)\ (1.6) = 10.1 \text{ ft packing}$$

Tower Loading, Flooding, and Pressure Drop

Assume 18-in. O.D. steel pipe, 16.8-in. I.D., crosssection area is 1.553 ft^2.

$\rho_L = 62.3\ lb/ft^3$

for 1 in. Bed saddles

Reference now to Figure 14-2 ID indicates that the tower would operate well below the loading zone at about 0.00636 (100)/.06 = 10.6% of the average loading condition. This is too low.

A new tower diameter can be assumed, or the limiting velocities can be calculated, and the diameter set from these. For illustration purposes, use the latter approach:

Average gas rate = (672) (1.553) = 1,047 lb/hr
Inlet gas rate = (701) (1.553) = 1,090 lb/hr
Average liquid rate = (695) (1.553) = 1,080 lb/hr
Outlet liquid rate = (724) (1.553) = 1,124 lb/hr

Because:

where V_g = superficial gas velocity, ft/sec

Solve for V_g: at abscissa = 0.0744

Read Figure 14-2 ID at flooding, ordinate = 0.150

Vg = 2.8 ft/sec flooding velocity

Read Figure 14-2 ID at upper loading, ordinate = 0.084

Vg = 2.1 ft/sec upper loading velocity

Base diameter on 50% of flooding

Operating velocity = 0.5 (2.8) = 1.4 ft/sec

Required tower crosssectional area at 50% of flooding:

If a standard 10-in. pipe is used (10.02-in. I.D.) the superficial

velocity = 0.897/0.546 ft^2 = 1.64 ft/sec.

% of flooding = 1.64 (100)/2.8 = 58.6%

% of upper loading = 1.64 (100)/2.1 = 78.3%

Pressure drop is approximately 0.5 in./ft packing height, Figure 14-21D.

This should be an acceptable operating condition, therefore, use tower diameter of 10-in. nominal Sch. 40 pipe. This pipe is satisfactory for 150 psig operating pressure (see Figure 14-88A and Figure 14-88B).

Correcting Height of Transfer Unit for Inert Gases in System

T (for 90°F) = 90 + 460 = 550°R at liquid inlet

T (for 90 + 72°) = 162 + 460 = 622°R max. at liquid outlet

M_A =17 for NH_3

M_B = 29 for air

P = 164.7/14.7 = 11.2 atm

V_A = 25-8 for NH_3 from Table 14-47, (molecular volume) Note: The value of Va is from Ref. 63. Molecular volume values vary between references.

V_B = 29.9 for air

Diffusion coefficient of NH_3 through air:

Diffusion coefficient of NH_3 through 3:1 H_2-N_2 mixture gas:

M_A = 17 for NH_3

M_B = 11.2 for inert gas mixtures, less NH_3
V_A = 25-8 for NH_3
V_B = (0.75) (14.3) + (0.25) (31.2) = 18.5 for H_2-N_2 mixture

Diffusion of ammonia through hydrogen–nitrogen inert gas:

For an ammonia–air system using mass rates in the tower the same as the H_2–N_2 system: G' = 1044/0.546 = 1912 lb/hr (ft^2) and L'= 1080/0.546 = 1980 lb/hr (ft^2)

From Figure 14-87 Extrapolated H_{OG} = 0.95 (this should be the maximum to expect)

Then substituting, for ammonia–air mixture:

Converting this to ammonia–H_2–N_2 mixture:

Then, substituting,

H_{OG} for this mixture

Tower Packing Height for Ammonia–Hydrogen–Nitrogen Mixture

Z = (1-929) (6.33) = 12.2 ft packing

For 1-in. Berl saddles allow 2.0 ft for good liquid distribution through the packing from the top.

Minimum packed bed depth = 12.2 + 2.0 = 14.2 ft

Use: 15 ft (or perhaps 18 ft to allow for variations in performance)

Expected Pressure Drop

Packing:	(15) (0.5 in/ft)	= 7.5 in.
Support:	(Grating type), approx.	= 1.0
Total drop		= 8.5 in. water

Inlet Liquid

For this small diameter tower, bring water into center of top of packing with a turned down 90° well, placed 6.9 in. above packing.

14.36. Mass Transfer With Chemical Reaction

Many absorption processes involve some chemical reaction; however, it is fortunate that satisfactory correlation can be made without delving into the complexities of the reaction.

In many instances the gas being absorbed is to be recovered from the solution, and hence the effluent from the absorption must be treated. Often the application of heat will release the gas, e.g., CO_2 from ethanolamine. In some cases, however, chemical treatment must be used. In order to fully use the absorbent liquid, recycle systems must be established, and the economics of absorbent selection become of

considerable importance; determining the size of the system in many cases. Because the absorption ability of the various combinations of gas and liquid vary rather widely, and depend upon system conditions, it is often necessary to examine more than one absorbent for a given gas. The performance of several important systems is given in summary or reference form:

14.36.1. Carbon Dioxide or Sulfur Dioxide in Alkaline Solutions

(14-126)

Where $K_G a$ = gas coefficient, lb mol/(hr) (ft^3) (atm)
n = CO_2 (or SO_2) absorbed, lb mol/hr
A = tower crosssection area, ft^2
Z = height of packed section in tower, ft
Δplm = log mean partial pressure of gas in inlet and exit gas streams, atm
f_a = fraction effective packing wetted

Design Procedure for Alkali-Absorbers

1. Calculate material balance, determining quantity of gas to be absorbed, and alkali required.
2. Estimate a tower diameter and establish its operating point from Figure 14-21. For the initial trial, set operating point at 50–60% of flood point system based on varying gas mass velocity at variable tower diameters for a fixed

3. Estimate the effective fractional wetted packing area by Figure 14-43 and Figure 14-66. As a general rule, try not to accept design if fraction wetted is less than 0.5. If it is less, adjust tower

conditions to raise value, bearing in mind that this factor is based on scattered and very incomplete data.

4. Determine $K_g a$ based on inlet alkali normality and percent conversion to carbonate (for carbon dioxide).

$K_g a$ Data and Corrections [42, 69, 72]

The necessary data required to properly design a CO_2 scrubber are separated into three sections, depending upon the CO_2 concentration in the feed gas. Corrections are necessary to convert "procedure based" information to the specific design basis of a given problem. The corrections outlined below are specific to the CO_2 concentration range being presented in the section.

Because there are insufficient data to serve completely for all types and sizes of packing, it may be necessary to estimate $K_g a$ values by ratioing packing surface areas and making the other appropriate correction for the problem conditions.

1. Inlet CO_2 concentration, 100 to 1,000 ppm by volume $K_G a$: Use Figures 14-93 A–E and Figure 14-94.

   a. Increases with increase in liquid rate, L, to the 0.2 power of the ratio,

   b. Increases with increase in gas rate, G, to the 0.35 power of the rate ratio, up to a rate of 500 lb/hr (ft^2). Above this rate the increase is reduced, being to the 0.15 power at rates near 1000 lb/hr (ft^2).

   c. Increases with increase in temperature to the 6.0 power of

the absolute temperature ratio.

d. Decreases with increase in pressure to the 0.5 power of the absolute pressure ratio.

Figure 14-93

(A) CO_2 absorption from atmosphere; effect of liquor rate on K_Ga at atmospheric pressure. Reproduced by permission of the American Institute of Chemical Engineers, Spector, N. A., and Dodge, B. F., *Trans. A.I.Ch.E.*, V. 42 © (1946) p. 827; all rights reserved. (B) CO_2 absorption from atmosphere; effect of liquor rate on K_Ga at atmospheric pressure and various gas rates. Reproduced by permission of the American Institute of Chemical Engineers, Spector, N. A., and Dodge, B. F., *Trans. A.I.Ch.E.*, V. 42 © (1946) p. 827; all rights reserved. (C) CO_2 absorption from atmosphere; effect of gas rate on K_Ga at atmospheric pressure. Reproduced by permission of the American Institute of Chemical Engineers, Spector, N. A., and Dodge, B. F., *Trans. A.I.Ch.E.*, V. 42 © (1946) p. 827; all rights reserved. (D) CO_2 absorption from atmosphere; effect of flow rates on K_Ga at elevated pressure. (E) CO_2 absorption from atmosphere; effect of tower pressure on K_Ga at various liquor rates.

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Figure 14-94

CO_2 absorption; effect of carbonate on K_Ga.

Reproduced by permission of the American Institute of Chemical Engineers, Leva, M., *A.I.Ch.E. Jour.*, V. 1 © (1955) p. 224; all rights reserved.

2. Inlet CO_2 concentration 1.5 to 5.0 mol% K_Ga: Use Figure 14-94Figure 14-95 and Figure 14-96 A-F.

a. Increases with increase in liquid rate, L, to the 0.28 power of the rate ratio,

b. No correction to gas rate, G, below loading. Above loading point, K_Ga increases with increase in gas rate G, to the 0.3 power of the rate ratio.

above loading

c. Effect of temperature and pressure same as section "1" above.

Figure 14-95

CO_2 absorption in sodium hydroxide (NaOH).

Used by permission of Leva, M., *Tower Packings and Packed Tower Design*, 2nd Ed., U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.) (1953).

Figure 14-96

(A) CO_2 absorption; effect of liquor rate. Reproduced by permission of the American Institute of Chemical Engineers, Tepe, J. B., and Dodge, B. E, *Trans. A.I.Ch.E.*, V. 39 © (1943) p. 255; all rights reserved. (B) CO_2 absorption; 1-in. packing data at constant gas rate. Reproduced by permission of the American Institute of Chemical Engineers, Leva, M., *A.I.Ch.E. Jour.*, V. 1 © (1955) p. 224; all rights reserved. (C) CO_2 absorption; 1-in. packing data at constant liquid rate. Reproduced by permission of the American Institute of Chemical Engineers, Leva, M., *A.I.Ch.E. Jour.*, V. 1, No. 2 © (1955) p. 224; all rights reserved. (D) K_Ga versus liquid rate for 4% sodium hydroxide (NaOH). Compiled from References 20 and 35. (E) ¾-in. packing data for system carbon dioxide-sodium hydroxide, gas rate constant.

Reproduced by permission of the American Institute of Chemical Engineers, Leva, M., *A.I.Ch.E. Jour.*, V. 1, No. 2 © (1955) p. 224; all rights reserved. (F) ½-in. packing data for system carbon dioxide-sodium hydroxide, gas rate constant. Reproduced by permission of the American Institute of Chemical Engineers, Leva, M., *A.I.Ch.E. Jour.*, V. 1, No. 2 © (1955) p. 224; all rights reserved.

2. Inlet CO_2 concentration 5.0 to 10.0 mol% K_Ga: Use Figure 14-94Figure 14-95 and Figure 14-96 A-F.

a. Liquor rate correction follows section “2” above.

b. Increases with increase in gas rate, G, to the 0.15 power of rate ratio.

c. Effect of temperature and pressure same as section “1.”

d. Calculate height of packing required:

Calculate tower pressure drop from Figure 14-21 for packing, and Figure 14-37Figure 14-38Figure 14-39Figure 14-40 and Figure 14-41 for support and grids.
e. Make specification sheet.

Example 14-15. Design a Packed Tower Using Caustic to Remove Carbon Dioxide from a Vent Stream

A process stream containing mostly nitrogen and carbon dioxide is to be scrubbed with 10% (wt) NaOH for CO_2 removal, but not recovery. The requirements are:

Inlet gas: CO_2 = 40.6 mol/hr
Inerts = 365.4 mol/hr
Avg mol wt = 20
Temperature = 90°F
Pressure = 35 psig = 49.7 psia
Outlet gas: $C0_2$ = 0.1 mol%
Inerts = 99.9 mol%
Max. allowable pressure drop = 2 psi
Liquid sp gr = 1.21

Solution

From a material balance; the NaOH required based on 25% conversion to Na_2C0_3:

$$2\ NaOH + CO_2 = Na_2CO_s + H_2O$$

lb 10% NaOH required/hr =

Assume 3.5 ft I.D. Tower, use 1½ in. steel Raschig rings. At operating point

For Figure 14-2ID

At flooding, ordinate = 0.0275, for abscissa = 0.752

At lower loading, ordinate = 0.012

Operating point = (0.012/0.0275) (100) = 43.8% of flood

This is only slightly lower than good practice, but because the operating point is at the lower loading point, continue with this selection. A smaller diameter tower might prove a better choice on further calculation. However, extreme caution must be used in designing too close to limits with packed towers. Very little of the data are exact, and often the range is not known.

Effective Wetted Packing

From Figure 14-43.

f_a = approx. 0.50 for ceramic material at L'/G' = 15.9

Because steel rings have thinner walls, it seems that the liquid should flow and wet this material a little better than ceramic.

Use $f_a = 0.6$

K_Ga from Figure 14-96 D uncorrected:

$(Kca)_{25} = 1.9$ lb. mol/hr (ft^3) (atm)

The conditions for K_Ga from Figure 14-96 D are 4 wt% NaOH, which is 1 normal, and $G' = 450$ lb/hr (ft^2), $L' = 2{,}000$ lb/hr (ft^2), pressure = 14.7 psia. The $(K_Ga)_{25}$ is also for 25% conversion to carbonate.

Correct the K_Ga for pressure, normality of NaOH, and liquid rate:

There is no correction for a gas rate at or below the loading point. Note that from Figure 14-94 the ratio of K_Ga at 1 normal and 2.5 normal solutions and for 25% conversion = 2.6/2.1.

Note: Because the solution temperature of data for Figure 14-96 D is 75–81°F, it is assumed ambient, and no correction is made. If the operating temperature were higher or lower, then a temperature correction multiplier should be used above, with the absolute temperatures ratio being raised to the 6.0 power.

Log mean driving force, for CO_2:

Inlet = 10%

Outlet = 0.1%

Packing Height

Experience in caustic-CO_2-steel Raschig ring systems indicates that the packing must be wetted better than the value of 60% (0.6); therefore a value of 0.85 is suggested.

then: Revised Z = 44.0 (0.6/0.85) = 31.1 ft

Allowance for distribution at top = 2 ft

Allowance for redistribution at mid-point = (2 ft) (2 sections) = 4 ft

Total packing height required = 31.1 + 2 + 4 = 37.1

Use 40 ft of packing.

Pressure Drop

From Figure 14-21C

ΔP = 0.38 in. H_2O/ft of packing

Total ΔP = (40) (0.38) = 15.2 in. H_2O

This is an acceptable figure

Arrangement

There are no data available on the liquid flow distribution vs. height for 1½-in. Raschig rings. Some information indicates that for 2-in. rings about 33% of the liquid is on the wall of a large tower after flowing through 20 ft of packed height, starting with good top liquid distribution.

To ensure *good* tower performance, use three 13 + ft packed sections of the 1½. Raschig rings. Two sections of 20 ft of packing would also probably perform satisfactorily, and be less expensive.

If Pall rings had been used, only two packed sections would be considered, because the general liquid distribution pattern is better. This would require a re-evaluation of the performance, and a probable reduction in total packed height.

a/ε^2	1.5 in. Metal Pall Rings 25	1.5 Metal Raschig Rings 110
Abscissa of Figure 14.21C	0.00272 (by ratio)	0.012
% Flood		43.8

On the basis of this better performance of the Pall ring, a *smaller* diameter tower *must* be selected and the tower re-evaluated for the new mass flow rates with this packing. Economics require that the higher packing cost, smaller tower diameter, new total packing volume, and tower pressure drop be considered.

14.36.2. NH_3–Air–H_2O System

H_{OG} data of Fellinger [27] Figure 14-87. Koa data of Dwyer and Dodge [21].

Carbon Raschig rings:

Size, in.	γ	α	τ	b	j
½	0.0065	0.90	0.65	0.39	0.310
1	0.036	0.77	0.78	0.20	0.103
1½	0.014	0.72	0.78	0.38	0.093

(For average temperature of 85°F, $H' = 2.74$ (lb mol/ft^3)/atm in dilute NH_3 solution, $D = 9.8 \times 10^{-5}$, $\mu = 1.97$)

Effect of humidity of entering gas found to be minor.

K_Ga increases as the 0.45 power of packing superficial area, and decreases with increase in temperature °F as the 0.635 power of inverse water temperature ratio.

where H' = Henry's Law constant, (lb mol/ft^3)/atm
G' = gas rate, lb/hr (ft^2)
L' = liquid rate, lb/hr (ft^2)

K_Ga and H_{OG} data of Wen [77], Figure 14-97 For ceramic Beri saddles:

(14-127A)

(14-127B)

where G_m = gas rate, lb mol/hr (ft^2)
L_m = liquid rate, lb mol/hr (ft^2)
For ceramic Intalox saddles:

(14-127C)

(14-127D)

Figure 14-97

Ammonia-air-water absorption data of Wen, with ceramic packing.

Used by permission of Chin-Yung, thesis, West Virginia University (1953).

Figure 14-98 shows effect of water temperature on $K_G a$ and H_{OG}.

Figure 14-98

Effect of water temperature on ammonia absorption.

Used by permission of Chin-Yung Wen, thesis, West Virginia University (1953).

14.36.3. SO_2-H_2O System (dilute gas)

$K_L a$ data of Whitney and Vivian [79]

Data for 1-in. ceramic Raschig rings, correlates reasonably well with 3-in. spiral tile and 1-in. coke.

(14-128)

lb mol/hr (ft^3) (lb mol/ft^3)

Temperature °F	b	H', lb mol/ft^3 (atm)
50	0.034	0.163
60	0.038	0.130
80	0.048	0.090
90	0.056	0.076

where K_La = overall absorption coefficient, lb mol/hr (ft^3) lb mol/ft^3

L = liquor rate, lb/hr (ft^2 of tower crosssection)

G′ = average gas rate of top and bottom of tower, lb/hr (ft^2 of tower crosssection)

H_{OL} = L′/p K_La, ft

ρ = liquid density, lb/ft^3

H_{OL} = height of overall transfer unit in terms of liquid film, ft

H′ = Henry's Law constant, c/p = lb mol/ft^3 (atm)

Effects of temperature on k_La:

$k_La \alpha e^{0.023T}$, T = °C

k_La is represented by the second term in the overall equation,

$k_La = b(L')^{0.82}$

k_La = Liquid film absorption coefficient, lb mol/hr (ft^3) lb mol/ft^3 = 1/hr

Effect of temperature on k_Ga = nil (assumed)

Reference 71 has excellent solubility data and absorption curves for the system.

14.36.4. Cl_2-H_2O System (for dilute gas concentrations)

Chlorine has limited solubility in water, Figure 14-99. KLa and H_{OL} data of Vivian and Whitney, [76] Figure 14-100. Data for 1-in.

ceramic Raschig rings.

(14-129A)

(14-129B)

(14-129C)

$K_L a \alpha T^6$ (of liquid, degrees absolute)
where T = absolute liquid temperature, °R
$K_L a$ = liquid coeff., lb mol/hr (ft^3) (lb mol/cu ft^3)
L′ = liquid rate, lb/hr (ft^2)
C_1 = concentration of chlorine in liquid, bottom of tower, lb mols/ft^3
C_2 = concentration of chlorine in liquid, top of tower, lb mols/ft^3

Figure 14-99

Solubility of chlorine in water.

Reproduced by permission of the American lnstitute of Chemical Engineers, Vivian, J. E. and Whitney, R. P., *Chemical Engineering Progress*, V 43 © (1947) p. 691; all rights reserved.

Figure 14-100

Effect of liquor rate on KL_a, and H_{OL} for chlorine and oxygen in water.

Reproduced by permission of the American Institute of Chemical Engineers, Vivian, J. E. and Whitney, R. P., *Chemical Engineering Progress*, V. 43 © (1947) p. 691; all rights reserved.

14.36.5. Air–Water System

The system is used in humidification and dehumidification. However, grid and slat packings are more commonly used types.

The H_{CG} data of Mehta and Parekh [49] is compiled by Leva [39, 62] for the ring and saddle packing.

Sherwood and Holloway [61] also studied the desorption of oxygen from water.

14.36.6. Hydrogen Chloride–Water System

The recovery of hydrogen chloride and the production of hydrochloric acid is performed effectively in adiabatic type absorption towers. Uncooled or adiabatic towers can be used to produce 33.5 wt% acid, and cooled towers will produce 35–36% acid with negligible vent losses [49] when using a feed gas containing 10–100% HCl. Due to the heat of absorption, heat dissipation must be taken care of by increasing the temperature of dilute acid product, or as a combination of this plus removal as water vapor in the vent. It is important to recognize that acid strengths greater than the constant boiling mixture are made in this tower from gases containing less than 20% hydrogen chloride. The lower the inlet feed gas temperature, the lower will be the acid product temperature.

Van Nuys [75] gives excellent thermodynamic data for HCl. Figure 14-101 gives the equilibrium for the 100% HCl gas feed in an adiabatic tower, and Table 14-48 summarizes performance for two concentrations of feed gas. From the data it can be seen that it requires fewer theoretical plates to make 32% acid from 10% feed gas than from 100% gas and at the same time yield a vent containing only 0.01 weight% HCl.

Figure 14-101

X-Y diagram for adiabatic absorption of HCI at 1 atmosphere.

Reproduced by permission of the American Institute of Chemical Engineers, Oldershaw, C. F., Simenson. L, Brown, T. and Radcliffe, *F. Trans. Sect. Chemical Engineering Progress*, V. 43, No. 7 © (1947) p. 371; all rights reserved.

Table 14-48 Adiabatic Absorption of HCl: Effect of Inert Gas

By permission, Oldershaw, C. F., Simenson, L., Brown, T., and Radcliffe, F. *Trans. Sect. Chem. Eng. Prog.*, 43, No. 7, p. 371 © 9 (1947), all rights reserved.

Operating Conditions Feed gas enters at equilibrium temperature at bottom of tower Make-up water enters at equilibrium temperature at top of tower Vent contains 0.1 wt.% HCl		
	Weight % in HCl in product	
Theoretical plates	**10 Mol % HCl feed**	**100 Mole % HCl feed**
2	26.5	20.5
4	32.0	30.5
6	33.8	32.8
10	34.7	34.0
∞	35.0	34.6

In general about 18 ft of 1-in. Karbate® or other acid resistant Raschig rings will satisfactorily absorb HCl from a gas stream up to 100% and produce 32% HCl acid.

Figure 14-102 illustrates a falling film type absorber using water jacketed tubes in the cooler-absorber. The tails tower removes last traces of HCI in the vents. Figure 14-103 is a preliminary selection chart for this type of unit.

Figure 14-102

Graphite type HCI absorption tower.

Used by permission of Falls industries, Inc.

Figure 14-103

Preliminary selection charts for HCI towers.

Used by permission of Falls Industries, Inc.

14.37. Distillation in Packed Towers

Packed towers are used in some distillation operations in preference to plate towers. Usually the selection requires an understanding of the fouling characteristics of the fluids of the system. These towers even

have been used in polymer forming operations. However, other contacting devices can be cleaned more easily. For some processes, the packed tower is much more effective as well as cheaper than a tray tower.

The more complicated separation for a three-phase system is discussed by Harrison [134].

14.38. Height Equivalent to a Theoretical Plate (HETP)

Distillation operations can best be expressed in terms of equilibrium relations and theoretical plates. Therefore, one of the correlating factors for various packings is the height of packing equivalent to a theoretical plate for the separation. Data for using this concept effectively is extremely meager, and contain many uncertainties as far as general application is concerned. For this reason the use of HETP is not popular. When good correlations are developed to predict HETP without test data, then this can be an effective means of expressing packing heights in distillation. Most HETP data have been obtained on small diameter (often laboratory size) columns using very small packings, and operated at essentially total reflux, or on moderate sized columns, but with limited systems. The scale-up of such data to industrial sizes is questionable. When the designer does not have actual data from similar services (often broad classes of similarity may have to be assumed to arrive at what might be termed a reasonable and safe value for HETP), then it can be helpful to contact manufacturer's technical service departments for their recommendations (which they normally will develop from application files). Keep in mind that the HETP is also unique to the packing size and configuration; therefore values obtained for one packing definitely do not accurately apply to another size or type. There is a rough relation, based on recognizing that large size packing requires greater HETP than small size, but pressure drop is greater for the small packing.

There can be a significant difference between the conventional "particle" packing, as represented in most of the illustrations of Figure 14-6 and the HETP values for most of the structured packing in Figure 14-6VFigure 14-6WFigure 14-6XFigure 14-6YFigure 14-6ZFigure 14-6AAFigure 14-6BBFigure 14-6CCFigure 14-6DDFigure 14-6EEFigure 14-6FFFigure 14-6GGFigure 14-6HHFigure 14-6IIFigure 14-6JJFigure 14-6KKFigure 14-6LLFigure 14-6MMFigure 14-6NN and Figure 14-6OO and 14-6YFF. These latter types (structured packing) offer HETP values varying from about 5 in. to 14 in., which are lower (greater efficiency) than the random particles which have values varying from 12 in. to as much as 36 in. Plots of HETP values for various packings have been presented in previous charts and discussions in this chapter.

Due to the unique HETP characteristics, it is important to consult the manufacturer on the specific system and operating conditions involved. Care must be exercised in selecting these, or any other type of packing, since plugging with suspended solids, polymer formation on surfaces, and similar mechanical problems can influence performance and life of the packing system.

Many correlations for HETP have been limited to Raschig rings or Berl saddles [25], both being the least efficient for mass transfer and pressure drop when compared to the more sophisticated designs represented in Figure 14-6. The guidelines given in a later paragraph are adequate for most of these applications.

Cornell, Knapp, and Fair [12, 13, 14] proposed the use of the transfer-unit concept for distillation, where:

(14-130)

where H_{og} = height of overall gas-phase transfer unit, ft
H_g = height of gas phase transfer unit, ft
H_l = height of a liquid-phase transfer unit, ft
m = slope of equilibrium curve
G = lb-mols gas/hour/ft^2
L = lb-mols liquid/hour/ft^2

where ϕ = correlation from Figure 14-103 and Figure 14-104 for a given packing

Figure 14-104

H_l correlation for various sizes of Raschig rings, Sc = Schmidt No. (N_{sc}).

Reproduced by permission of the American Institute of Chemical Engineers, Cornell, et al., *Chemical Engineering Progress*, V. 56, No. 8 © (1960) p. 68; all rights reserved.

Cp = correction factor for high gas rates, from Figure 14-105

Figure 14-105

H_l correlation for various sizes of Berl saddles.

Reproduced by permission of the American Institute of Chemical Engineers, Cornell, et al., *Chemical Engineering Progress*, V. 56, No. 8 © (1960) p. 68; all rights reserved.

μ_1 = liquid viscosity, lb/ft (hr)
ρ_l = liquid density, lb/ft^3
D_1 = liquid-diffusion coefficient, ft^2/hr
Z = height of packing, ft

For Raschig Rings:

(14-131)

For Berl Saddles:

(14-132)

where Ψ = parameter for a given packing, from Figure 14-106 and Figure 14-107

Figure 14-106

Liquid film correction factor for operation at high percent of flood.

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Figure 14-107

H_g, correlation for various sizes of Raschig rings.

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S_Cg = gas-phase Schmidt Number = $\mu_g/\rho_g D_g$
μ_g = gas viscosity, lb/ft (hr)
ρg = gas density, lb/ft^3
Dg = gaseous diffusion coefficient, ft^2/hr
D = column diameter, in.
Z = packed height, ft
$f_1 = (P_1\ 2.42)^{0.16}$
$f_2 = (62.4/Pi)^{1.25}$
$f_3 = (72.8/a)^{0.8}$
μ_1 = liquid viscosity, lb/ft (hr)
ρ_l = liquid density, lb/ft^3
σ = surface tension, dynes/cm

In Reference 14, the authors modified the equations for Hg and Hi as follows: (a) eliminate column diameter correction above 24 in., and (b) columns with *good* liquid distribution probably can allow elimination of the packing height correction.

Two separate investigators have evaluated the various correlation methods, and reported that the Cornell, *et al.* method is significantly better than others. These were 1-and 2-in. metal Pall rings and 1-in. ceramic Intalox saddles [24] and ¾-, 1½-, and 3-in. Raschig rings [68].

Whitt [80] has correlated literature data of commercial size Raschig ring packing, as shown in Figure 14-109. The range of plotted data is indicated, and the suggested design lines are good medians. The HTU values are for gas film controlling absorption systems, and the HETP data are for distillations at 760 mm Hg and below. These values should be usable for most pressure systems. The viscosity of the liquid ranged 0.35 to 1.0 centipoise. The equation for the HETP line is [34]:

(14-133)

and for HTU_G

(14-134)

Figure 14-109

HETP and HTU correlation for tower random packings.

Used by permission of Whitt, F. R., *British Chemical Engineering*, July (1955) p. 365; all rights reserved.

In general, for the same liquid and vapor rates, the HETP and HTU values for Berl saddles and others with a/ε^3 lower than Raschig rings should be lower. Correlating data is not available, except the tests of Teller [71] which indicate 1-in. Berl saddles have HETP values of 0.75 compared with 0.85 for 1-in. Raschig rings.

The accuracy of establishing HETP values for new conditions in the same packing referenced to known values, as suggested by Planovski [57], has not been tested by other literature references.

(14-135)

∗represents known conditions
P = system pressure, absolute

Figure 14-108.

H_g, correlation for various sizes of Berl saddles.

The operation of packed towers under vacuum conditions is not well defined in the literature. However, the work of Hands and Whitt [33] specifically evaluates several systems operating from 20 to 760 mm Hg abs. Their recommended limiting vapor velocity is:

(14-136)

or

(14-137)

Operating values are recommended to be two-thirds of the limiting values. From the data, it appears that entrainment (internal may be more significant than external) becomes a limiting factor before the

flooding values predicted by Figure 14-21C and Figure 14-21D.

Eckert [121, 122] and others have examined data, and presented considerable tabulations of HETP and HTU values for random packing in various organic-water, polar organic, and hydrocarbon systems. Some of these are now obsolete, due to the older type/style packing tested; however, the results still have value as a base reference if it is recognized that the newer types/styles are more efficient, giving lower HETP and HTU values (the structured packings are discussed in a separate section of this chapter).

14.39. HETP Guide Lines

For industrial process equipment, some general guide lines in this undefined area are (for particle packing):

1. Never use HETP less than 12 in. if tower is 12 in. dia. or larger; for general assumption, use HETP = 1.5 to 2.0 ft.
2. Use HETP = H_{OG} or H_{OL} if other data are not available
3. Use HETP = Column diameter (over 12 in. dia.) if no other information is available, up to 48 in. dia.

HETP values appear to vary somewhat within the distillation system, while certainly varying with the size and style of packing. In general the larger, more open packing designs exhibit higher HETP values; while the smaller particle type packing and the compact styles exhibit significantly lower values for the same system. A brief guide to recently published values for various systems would be helpful in establishing the right order of magnitude for a system in design.

Kister [90] has tabulated published HETP data for a wide variety of process systems, and proposes that using good experimental data to interpret for other systems can be just as effective (accurate) as using values calculated using most mass transfer techniques.

Influences on HETP Values

Various authors [90] and researchers have published details of factors that influence detail of HETP values. The lower the HETP value (i.e., number of feet (or inches) per theoretical plate/tray/stage) the more efficient is the packing (random or structured) for any particular separation system. Other influences include:

1. Packing surface area/unit packing volume increases, resulting in more efficient packing performance (lower HETP). For structured packing the more narrow passage-way between sheet components results in more efficient performance [90].
2. Uniformity of packing surface on a specific random element of packing [90]
3. Uniformity of liquid (vapor) distribution increases packing performance efficiency.
4. Liquid and vapor loadings have little effect on HETP for random packing up, to the point between loading and flooding.
5. For structured packings, as loadings increase HETP increases. This is more pronounced for wire-mesh types, and the effect is less for the corrugated sheet packings [90].
6. Pressures above 1 or 2 psia have small effects on HETP for random and structured packings. For high vacuum, the data are not totally firm as to the consistency of effects; generally HETP increases (efficiency becomes lower). At pressures above 200–300 psia structured packing HETP increases (lower efficiency) [90].

14.40. Transfer Unit

The transfer unit concept is also applicable to distillation in packed towers. Height of the packing required is:

(14-138)

For usual applications in the rectifying section where the number of transfer units N_{OG} is given by:

(14-139)

for the stripping section, usually:

(14-140)

The height of the transfer unit has not been satisfactorily correlated for a wide variety of systems. If pilot plant or other acceptable data are available to represent the system, then the height of packing can safely be scaled-up to commercial units. If such data are not available, rough approximations may be made by determining Hg and Hl as for absorption, and combining to obtain an H_{OG} (Ref. 74). This is only very approximate. In fact it is due to obtained the lack of any volume of data obtained from commercial units that many potential applications of packed towers are infact designed as tray towers.

(14-141)

Based on the two resistance theory [127]

(14-142)

If, over a single stage, the equilibrium curve and the operating lines

are assumed straight to be, then [127]:

(14-143)

(14-144)

L_M/G_M is the internal reflux ratio that is the slope of the operating line, in either the rectifying or stripping sections. The value of L_M/G_M usually will not vary significantly from 1.0. $\lambda = mG_M/L_M$ is the ratio of the two slopes and remains close to 1.0 for different separation systems [127]. For these conditions HETP $\cong$ $(HTU)_{OG}$ and HETP become an important design variable. For $\alpha >> 1.0$ for easy separations and λ is less than or much greater than 1.0. See Figure 14-110 and Figure 14-111, which illustrate the effect on HETP of under-irrigating the bed by the distributor.

Figure 14-110

Vapor-liquid equilibrium showing λ and application cases referred to in the text.

Used by permission, Koshy, T. D., and Rukovena, F. Jr., *Hydrocarbon Processing* V. 65, No. 5 (1986) p. 64; all rights reserved.

Figure 14-111

Effect of under-irrigating the wall area of a column using random packing. Note: % blanked refers to packing crosssectional area not irrigated. At 0% blanked the best HETP is found. To the right of 0 towards the wall (50% blanked) the HETP becomes poorer until the wall is reached. For the center section (to the left) of the tower from the 0% to 50% of the center area of tower area blanked, the poorer HETP is again found.

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A minimum HETP or HTU represents a maximum separation efficiency, with α representing the relative volatility, i.e., vapor and liquid phase compositions of the more volatile component in a binary system:

The following examples are presented here by permission of the authors [127] and *Hydrocarbon Processing:*

Case I: γ is much less than 1. The circumstances when this happens are [190]:

- High purity of the more volatile component, i.e., $m \rightarrow \uparrow/\alpha$.
- High reflux ratio, i.e., $L_M/G_m \rightarrow 1.0$.
- Both foregoing conditions mean $\lambda \rightarrow 1/\alpha$.

Case II. $\lambda = 1$. The following situations will produce this condition [127]:

• Very low reflux ratio for high purity rectification, i.e., $x \rightarrow 1.0$.
• Very high reflux ratio for high purity stripping, i.e., $x \rightarrow 0$.
• Total reflux for a symmetric separation. Note, the term "symmetric separation" is used here to mean that on a McCabeThiele diagram, the liquid phase compositions of the overhead product and bottom product are roughly equidistant from 0.5.

Case III. λ is much greater than 1. The circumstances when this occurs are [127]:

• High purity of bottoms products, i.e., $x \rightarrow 0$ and $m \rightarrow \alpha$
• Low L/V approaching total reflux, i.e., $L/V \rightarrow 1.0$
• Both foregoing conditions mean that $\lambda \rightarrow \alpha$

where G_M = gas flow rate, kg-mols/hr, or lb mols/hr
HETP = height equivalent to a theoretical plate, m, or ft
HTU = height of a transfer unit, m or ft
L_M = liquid rate, kg-mols/h, or lb mols/hr
m = slope of the equilibrium line
x = mol fraction of the more volatile component in the liquid phase
y = mol fraction of the more volatile component in the vapor phase

Greek symbols

α = relative volatility between components 1 and 2
$\lambda = m\ G_M/L_M$, ratio of the slopes of equilibrium and operating lines

Subscripts

G = gas phase
L = liquid phase
OG = overall gas phase

Packed column performance can use either the HETP or HTU concepts; the HTU is somewhat more complicated but no more correct. HETP, adapts itself to direct use from tray-by-tray digital computer calculations, and is thereby a little more direct.

The packed column has been quite useful in distillation, stripping, and absorptions processes and has become competitive with many types of distillation tray designs or types/styles.

Bolles and Fair [29] present an analysis of considerable data in developing a mass-transfer model for packed tower design; however, there is too much detail to present here.

Example 14-16: Transfer Units in Distillation

A benzene-toluene mixture is to be separated in a tower packed with 1-in. Berl saddles. The feed is 55.2 mol% (liquid feed, saturated), and an overhead of 90 mol% benzene, and bottoms of not more than 24 mol% benzene is desired. Using the data of Ref. 51, plotted in Figure 14-112, determine the number of transfer units in the rectifying and stripping sections using a reflux ratio (reflux to product, L/D) of 1.35.

Figure 14-112

Vapor-liquid equilibrium (data only), benzene-toluene. Diagram notes for this text by this author.

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Solution

Referring to Figure 14-112 for the graphical solution:

Rectifying section operating line slope =

Note that point 7 can be determined by the intersection of the rectifying operating line and the feed condition line 8–7.

Establish the location of the feed, bottoms and overhead compositions on the graph. Draw in the operating lines as for a distillation in a tray column.

To establish the transfer units draw in line A–B–C so that it is always half-way vertically between the equilibrium line and the operating line, making dimension 1–2 equal to 2–3. Begin drawing the transfer units at the overhead product 4, such that 4–9 equals 9–5, then drop vertically to the operating line and repeat the process, always making the line A–B–C bisect the horizontal portion of the step. At the feed point, re-start the stepwise process if the transfer unit step does not terminate at the feed point 7.

For this example, the number of transfer units is:

Rectifying:	2.9±
Stripping:	2.6±
Total	5.5± units

The reboiler for the column is in addition to this; however, the bottoms were specified as being the inlet from the reboiler. For most purposes the reboiler can be considered to be one additional transfer unit.

Alternative

An alternative method to determining the number of transfer units is the graphical integration of dy/(y∗ – y). The procedure is basically the same as for absorbers, that is:

1. For assumed values of x (mol fraction of component under consideration in liquid) from bottoms to overhead, read values of y (vapor under operating conditions corresponding to x) and values of y∗ (vapor in equilibrium with x) from the equilibrium line.
2. Calculate l/(y∗ – y) at each selected point, thus:

x	y∗	y	y∗ – y	l/(y∗ – y)
0.24	0.44	0.24	0.20	5.0
0.30	0.52	0.328	0.192	5.21

3. Plot y, from y bottoms to y overhead versus 1/(y∗ – y). The position of y-feed can be noted on the graph, and the integration so arranged as to reveal the split between rectifying and stripping transfer units. The total number by this method should check closely with the graphical stepwise method.

Height of Transfer Unit

The height of a transfer unit for this system is not available, therefore it may be *roughly* approximated by the method of additive H_G and H_L which is questionable at best, or the approximation of 2 ft for HTU may be used. The latter is just about as reliable as the former.

Then: height of tower packing, using 1-in. Berl saddles:

From transfer unit = (2) (5.5) =	11 ft
Allowances: for top distribution:	2 ft
Sub-total	13 ft

Extra, 20%	2.5 ft
Total Z,	15.5 ft
	use 16 ft

The tower must be designed for throughput—diameter determined and pressure drop established.

Check Theoretical Plate Basis

To determine HETP by approximate method, see Table 14-49 for benzene-toluene:

HETP = 1.0 to 1.5 ft

Select HETP = 1.25 ft

Safety factor suggested = 2, in any case a value not less than 1.25

Table 14-49 HETP Estimates for Distillation Applications

Data for table compiled from respective manufacturer's published literature.

Contact manufacturers for specific design recommendations			
System	**Pressure Range If Known**	**General Packing Type/Style/Make**	**Estimating HETP* Ft or In. Marked**
Iso-octane/toluene	100mm Hg	Hy-Pak No. 2	2.0–2.7
Same	740mm Hg	Hy-Pak No. 1	0.7–2.7
Para/ortho xylene	740mm Hg	Metal Intalox No. 25	0.8–1.3
Same	740mm Hg	Meta! Intalox No. 40	1.3–1.55
Same	740mm Hg	Metal Intalox No. 50	1.75–2.15
Chlorinated HC	Vacuum	Metal Intalox No. 25	2
Chlorinated HC	Vacuum	Metal Intalox No. 40	2.4

Chlorinated HC	Vacuum	Metal Intalox No. 50	3.5
Iso-octane/toluene	740mm Hg	Pall Ring, 1 in. Metal	1.0–2.0
Iso-octane/toluene	740mm Hg	Pall Ring, 1½ in. Metal	0.75–1.0 (3.5) **
Iso-octane/toluene	740mm Hg	Pall Ring, 2 in. Metal	1.5–2.2
Methanol/water	740mm Hg	Pall Ring, 1 in. Metal	0.65–0.8 (1.2) **
Isopropanol/water	740mm Hg	Pall Ring, 1 in. Metal	0.6–1.5
Benzene/toluene	740mm Hg	Pall Ring, 1 in. Metal	1.0–1.5
Acetone/water	740mm Hg	Pall Ring, 1 in. Metal	0.9–1.2 (1.4) **
Same	—	Flexirings, 1 in. Metal	1.6–1.8 (2.3) **
Same	—	Flexirings, 2 in. Metal	1.8–2.2 (2.4) **
Same	—	Koch Sulzer Metal	0.45–0.9
Light hydrocarbon	400 psia	Goodloe Metal	approx. 0.75
Propane/butane	235 psia	Goodloe Metal	approx. 0.80
Chlorobenzene/ethylbenzene	50mm Hg	Montz structured metal	5 in.–17 in.
Chlorohexane/n-heptane	1 atm	#2 Nutter Ring	22 in.–30 in.
Chlorohexane/n-heptane	5 psia	#2 Nutter Ring	25 in.–30 in.
Various sys.	vacuum	Goodloe metal, various	5 in.–8 in.
Iso-Octane/toluene	1 atm	Cascade Miniring, #3	22 in.–28 in.
Iso-Octane/toluene	1 atm	Cascade Miniring, #2	18 in.–24 in.

h-naphtha/light gas oil	unknown	Gempak, ½ in. crimp	13 in.–15 in.
h-naphtha/light gas oil	unknown	Gampak, 1 in. crimp	22 in.–27 in.
h-naphtha/light gas oil	unknown	Gempak, ¼ in. crimp	8 in.–10 in.
Ethylene dichloride/benzene	1 atm	ACS-X Mesh	4 in.–9 in.
Methylcyclohexane/toluene	1 atm	ACS-X-200 Mesh	3.5 in.–12 in.
Unknown	unknown	Koch structured Flexipac	17 in.
General/average	unknown	Koch/Sulzer (R) structured	3 in.–9 in.
Ortho/para-	16mm Hg abs	Koch/Sulzer (R) structured	5 in.–16 in.
Ortho/para-	100mm Hg abs	Koch/Sulzer (R) structured	4.5 in.–8 in.
*Based on industrial data or commercial sized tests, note some values in inches.			
**At very low gas rates.			

Therefore use:

HETP = 24 in. = 2 ft

From Figure 14-112 it is evident that the number of theoretical plates and number of transfer units are not the same. When stepped off, the number of theoretical plates is 6+.

Height of packing = (6) (2) =	12 ft
Allowance for distribution =	2 ft
Total	14 ft

Use: 14-ft packing, 1-in. Berl saddles.

Because the 16 ft of packing by the HTU method is larger, this would be the recommended safe height to use.

For comparison, note the relative increase in the number of transfer units if the operation were at a higher pressure as shown by the dotted line for 500 psig.

Strigle [39] discusses packed column efficiency (HETP) in considerable detail. Most of his published data refers to work of Norton Chemical Process Products Corp.

A Norton [39] correlation for modern, random dumped packings used for distillation up to 200 psia is (use high performance internal distributors and supports) from surface tension of 4 dynes/cm but less than 33, and liquid viscosity of at least 0.83 to 0.08 cPs but not greater:

(14-145)

Values of n in the equation for selected specific packings are given in Table 14-50 for random packing and Table 14-51 for structured packing.

Table 14-50: Constant n for HETP Correlation*, Random Packing.

Used by permission from Strigle, R. E., Jr., *Packed Tower Design and Applications: Random and Structured Packings;* 2nd Ed. Gulf Pub. Co. (1994).

Tower packing	Value of n
#25 IMTP® packing	1.13080
#40 IMTP® packing	1.37030

#50 IMTP® packing	1.56860
1 in. Pall Ring	1.13080
1½ in. Pall Ring	1.39510
2 in. Pall Ring	1.65840
1 in. Intalox® Saddle	1.13080
1½ in. Intalox® Saddle	1.41570
2 in. Intalox® Saddle	1.72330
**IMTP and Intalox are registered names of Norton Chemical Process Products Corp.	
*Use with Equation 14-125.	

Table 14-51 Constant n for HETP Correlation* for Intalox Structured Packing. Used by permission from Strigle, R. E., Jr., *Packed Tower Design and Applications: Random and Structured Packings;* 2nd Ed. Gulf Pub. Co. (1994).

Packing size	Value of n
1T	0.76830
2T	1.01280
3T	1.38680
**IMTP and Intalox are registered names of Norton Chemical Process Products Corp.	
*Use with Equation 14-125.	

Strigle presents typical separation efficiency ranges for Intalox® metal tower packing, for systems with relative volatility not greater than 2.0.

Packing size	HETP (ft)
#25	1.2 to 1.6
#40	1.5 to 2.0
#50	1.8 to 2.4

where σ = surface tension, dynes/cm
μ = liquid viscosity, centipoise
n = constant for HETP Equation 14-145

Data for table compiled from respective manufacturer's published literature.

The summary of HETP values of Vital [192] for various types and sizes of packings are believed to be referenced to typical industrial distributors for the liquid. This variation can influence the value of HETP in any tabulation; the effect of distributor design is discussed in an earlier section of this chapter. Porter and Jenkins [143] developed a model to improve the earlier models of Bolles and Fair from about 25% deviation to about a 95% confidence using a 20% factor of safety [139].

Strigle [139] recommends: (See reference for related details):

1. For easy separations (less than 10 theoretical stages) a 20% design safety factor can be applied to a typical HETP value.
2. For separations of 15 to 25 theoretical stages a 16% design safety factor should be applied to the HETP.
3. For very difficult separations, the design HETP should be carefully evaluated by calculation and actual data when available.
4. HETP values for random dumped packing have been found to be 25% greater at a greater viscosity than a lower viscosity, i.e., viscosity change from 0.15 cPs to 0.44 cPs.

14.41. Cooling Water With Air

Wood-or plastic-filled towers for cooling water by using air are quite economical for certain heat loads and geographical locations. The costs of installation and operation must be compared with once-

through water costs at any location to arrive at a properly based decision. The four commercial tower types are:

14.42. Atmospheric

This tower depends upon atmospheric airflow to blow horizontally (or nearly so) through the tower (Figure 14-113). These towers must be in relatively open areas to receive the available wind from any particular direction. Wind velocities of 4.5 to 6.5 mph are necessary for reasonable operation. The towers operate with wind flowing across of the falling water and range from 30–55% effective. They are not capable of producing water at a temperature much closer than 4°F of the wet-bulb temperature entering air. They require no fan, but do consume power to pump the water to the (relatively high) top of the tower. Ground area requirements may be large.

Figure 14-113

Atmospheric cooling tower.

Used by permission of The Pritchard Corp. (now, Black and Veatch Pritchard).

The spray-filled tower, Figure 14-114, is also an atmospheric type, containing no fill other than the water sprays, and no fans. The water-air is achieve by the spray distribution system [144]. This design is often used where higher water temperatures are allowed, and in situations where excessive contaminants building up in the water would cause fouling of other direct contact heat transfer surfaces.

Figure 14-114

Atmospheric spray tower, air flow aspirated by pressure-spray water distribution system. Usually applied in small sizes.

Used by permission of Hensley, John C. (ed), *Cooling Tower Fundamentals* 2nd Ed. (1985), The Marley Cooling Tower Co., a United Dominion Company.

14.43. Natural Draft

This tower depends upon natural draft action in the same way as a chimney, to draw cool air in at the bottom and expel it out the top as warm, moist air (Figure 14-115). The action of the tower depends upon the atmospheric temperature, so on a hot day the action of the tower may be less than on a cool day. These towers are relatively large, and require power for pumping the water to a point in the tower which is usually lower than for an atmospheric tower. There are no fan costs. Units have been built 310 ft high, base diameter 210 ft and a throat of 120 ft, widening to 134 ft in diameter at the top [130].

Figure 14-115

Component parts of modern natural draft tower.

Used by permission of Hamon Cooling Towers, Inc.

14.44. Forced Draft

This type of tower uses fans at the base to force air through the tower fill or packing (Figure 14-116). Due to the relatively low outlet air

velocity, there is a tendency for discharged hot air to recirculate into the fan intake and reduce tower performance. The fan handles only atmospheric air thereby reducing its corrosion problem when compared to the fan on an induced draft tower. The tower size for the forced as well as the induced draft unit is considerably less than for an atmospheric or natural draft unit due to the higher heat transfer rates.

Figure 14-116

Crosssection of low-head forced-draft tower showing fan housing arrangement, filling, water distribution spray system and spray eliminators.

Used by permission of Foster Wheeler Corp., Cooling Tower Dept.

14.45. Induced Draft

This tower uses fans at the top of the tower to draw air in as its base through the fill and out of the fan discharge (Figure 14-117AFigure 14-117BFigure 14-118 and Figure 14-119). In this type of mechanical draft tower, the hot, moist air discharges vertically (usually) to the atmosphere with enough velocity to eliminate the possibility of recirculation of this air into the base of the tower. The moist air is corrosive to the fan parts, and they therefore require the protection of coated plastic or special metal blades, sealed motors and reduction gears.

Figure 14-117A Counterflow induced draft cooling tower.

Used by permission of The Pritchard Corp. (now, Black and Veatch Pritchard).

Figure 14-117B

Counterflow induced draft cooling tower for industry.

Used by permission of The Pritchard Corp., Bull 4.9080A (now, Black and Veatch Pritchard).

Figure 14-118

Spray-filled counterflow induced draft cooling tower.

Used by permission of The Marley Cooling Tower, a United Dominion Co.

Figure 14-119

Cross-flow induced draft cooling tower.

Used by permission of The Marley Cooling Tower, a United Dominion Co.

14.46. General Construction

Most cooling towers are built of redwood or cypress. However,

special conditions and atmospheres dictate other types of construction.

Materials:

1. Framework
   - Heart redwood
   - Cypress
   - Galvanized steel Brick or Concrete

2. Casing
   - Heart redwood
   - Cypress
   - Corrugated asbestos board, or some combination with redwood.
   - Brick or concrete block

3. Fill or Packing
   - Heart redwood
   - Cypress
   - Asbestos-cement boards or strips
   - Plastic sheet, grids or pieces

4. Drift Eliminators, same as 3.
5. Louvers, same as 3.
6. Miscellaneous Hardware
   - Monel, galvanized or other corrosion resistant metals

7. Fans and Drivers

Axial or propeller blade fans are either belt or gear driven. Some drivers are variable speed motors, and some fans have variable pitch blades. In special circumstances, steam turbine, gas or gasoline engine drivers are used. Gears should be carefully specified to avoid overload, and should be specially sealed to prevent moisture entering the case.

14.47. Cooling Tower Terminology

Wet Bulb Temperature

This is the temperature of air at which it would saturate without a change in its heat content. It is the theoretical minimum temperature to which water may be cooled in a cooling tower. In actuality this temperature can never be reached, but only approached. The selection of the proper wet bulb temperature is very critical, because it is the single most important factor in tower rating. The selected temperature should be high enough to include 95% of the maximum readings recorded during the time that is most critical or important in the cooling service.

If the temperature is too high, an expensive tower will be specified; and of course, if too low, the cooling load service will be perform less well during the times when the wet bulb exceeds the specified value. At constant inlet humidity and constant rates for liquid, L′, and air, G′, the effect of changing wet bulb on the performance factor Ka V/L′ is only 1.2% with no trend dependent on rates [38].

Approach

This is the temperature difference between the cold water outlet temperature and the wet bulb temperature of the air. The smaller the

approach the more difficult the cooling job, and the larger the required tower. For a fixed cold water temperature, changing the wet bulb temperature by one degree can make a significant difference in tower requirements. Usually an approach of 5°F is considered minimum.

Range

This is the temperature difference between the warm water flowing into the tower and the cold water flowing out. The range determines the heat load on the tower, which in turn reflects the requirements of the cooling water service. The average reduction in KaV/L′ for each 10°F increase in hot water inlet temperature is 2% [38].

Drift Loss or Windage Loss

This is the amount of water lost from a tower as fine droplets become entrained in the leaving air. For an atmospheric type tower this is usually 0.1–0.2% of the total water circulated. For mechanical draft towers it is usually less.

Make-up

This is the water required to be added to the system to make-up for losses by evaporation, drift loss, and blowdown.

Blowdown

This is the amount of water continuously or intermittently removed from the system, to maintain a predetermined water analysis with respect to chemicals and dissolved gases. The buildup of solid or chemical concentration that will occur with continued evaporation and no blowdown can become very corrosive and harmful to the metal and wood parts of the system. In addition, deposition of salts on exposed surfaces, and accumulation of sludge in the basin of the tower can influence performance as well as affect the life of the tower.

Recirculation

This is the portion of exit or outlet air from the tower that recirculates back to the inlet of the fresh air to the tower. To keep this low, it is

important to space towers away from each other, as well as from any structures which can deflect the moist exit air back to the inlet. Due to recirculation, the wet bulb temperature at the tower inlet may be different from that at a point 100 yards away. The recirculation of induced draft towers is usually less than for the forced draft type due to the upward velocity of discharge of the air.

Normal recirculation in average installations for forced draft may run at between 3–10% of total inlet air, and 1–8% for induced draft towers, all depending upon the location and wind conditions during any day or season. Some towers can be arranged to have less than 1% recirculation. If prevailing conditions are suspected of being conducive to recirculation, it should definitely be allowed for in local design of the tower. Recirculation increases the wet bulb temperature of entering air and increases the total air required (and hence size of all equipment) for a given tower performance.

14.48. Specifications

Specifications for performance rating are usually set by the process engineer, with the rating selection performed by the cooling tower manufacturer. Each manufacturer has packing arrangements with known, specific, performance characteristics, and has developed size modules for standard cells (usually 6 ft × 6 ft or 8 ft × 8 ft) together with the associated fan requirements. Some of this information is tabulated in the catalog literature. Specific economical ratings must consider the performance specified in light of the local application of the tower. To do otherwise can very often lead to excessive costs for this type of equipment. An informational specification sheet to be used by the process engineer is given in Figure 14-120.

Figure 14-120

Cooling tower specifications form.

Additional information is available from the Cooling Tower Institute, including ATF-105 *Acceptance Test Procedure for Water-Cooling Towers*, STD-101 *CTI Grades of Redwood Lumber* [31], STD-102 'Structural Design Data' [70], and TSC-302 *Cooling Tower Wood Maintenance* [16].

It is recommended that performance tests be specified and conducted in accordance with the Cooling Tower Institute procedure, as this gives the process engineer a standard of reference. Most cooling tower manufacturers are members of this Institute.

Manufacturer's bid proposals should include all of the information needed for the specification sheet and in addition, details of construction, details regarding driver, gear, etc., and a guaranteed performance curve showing the effect of ±10% change in water quantity and lower wet bulb temperature, similar to Figure 14-121.

Figure 14-121

Typical performance for design gpm.

Used by permission of Whitesell, J., *Chemical Engineering*, Jan. (1955) p. 187.

For rating by the manufacturer, the process engineer must specify and

consider:

1. Water rate, gpm

2. Inlet water temperature, °F
3. Outlet cold water temperature, °F
4. Design wet bulb temperature, °F, for the location of construction of the tower.
5. Water condition (sandy, oily, etc.) and type (river, canal, harbor, sea). Contaminating chemicals and/or minerals should be identified. Type of water treatment.
6. Drift loss or mist loss, usually maximum of 0.2% of design water flow rate.
7. External wind force or loading for standard design. Most designs are for 30 lb/ft^2, although specific geographical locations may require other specifications. Give minimum, and average wind velocities with compass direction for atmospheric and natural draft towers.
8. Geographical location; plant site location, general proximity to other structures and other factors relating to recirculation of exit air back to the inlet of the tower.
9. Type and specifications on fan driver, gear types, power voltage, phase, cycles. Motors should at least meet specifications equivalent to totally enclosed, fan cooled, or if in explosive hazardous area, TEFC Class I, Group D (except this not acceptable in hydrogen or acetylene atmosphere). Due to moisture conditions around this equipment, it should be protected against moisture penetration and corrosion.
10. Number, type, height, area requirements for cooling coils (if any) to be installed in tower basin by purchaser.
11. Power costs for fan and pump horsepower, approximate pump efficiency for water, and any special data peculiar to the economics of the installation. This will allow the manufacturer to select a tower on the wasis the economic factors involved.

12. Items to be furnished by the purchaser, such as concrete basin, anchor bolts, electrical components, external piping, material handling to job site. If the tower manufacturer is to perform a turn-key or package job, this should be specified, as in some instances the tower manufacturer may not be in a position to do this.
13. Fire protection if unit is to be allowed to stand dry for prolonged periods, or is required by insurance.

14.49. Performance

The cooling tower cools hot water with cool air by countercurrent (or cross-current) flow of the two fluids past each other in a tower filled with packing. This involves both mass and heat transfer. The water surface that exists on the tower packing is covered with an air film that is assumed to be saturated at the water temperature. The heat is transferred between this film and the main body of air by diffusion and convection. Detailed presentations of the development of cooling tower theory are given in References 39 and 46.

Figure 14-122 and Figure 14-123 indicate the variables in tower performance. Also see Reference 150.

Figure 14-122

Diagram of counter current air-water cooling tower.

Figure 14-123

Driving force diagram for cooling tower.

The packing or fill is arranged to prevent a droplet of water from falling the full height of the tower. As it falls it hits a packing member, splashes, forms a film, drops off and falls to hit the next packing member. The countercurrent stream of air sweeps across these drops and films to effectively cool the water and humidify the air. As the water flows down through the tower, its temperature may drop below the dry bulb temperature of the inlet air to the tower. It can never go below the inlet air wet bulb temperature, in fact, it just *approaches* this wet bulb. One of the controlling features in tower design and performance is how close these two temperatures, inlet air wet bulb and outlet water, are expected to operate. The driving force for the cooling is the difference in enthalpies of the film of air surrounding the water and that of the main body of the air.

The number of transfer units or tower characteristic is based on overall heat and mass transfer:

(14-146)

where h′ = enthalpy of saturated air film at bulk water temperature, Btu/lb
h = enthalpy of the main air stream, Btu/lb
t_1 = entering warm water temperature, °F, at top of tower
t_2 = outlet cool water temperature, °F, bottom of tower
t = bulk water temperature, °F

K = mass transfer coefficient, lb water/hr/ft^2
a = contact area, ft^2/ft^3, tower volume
V = active cooling volume, ft^2/ft^3, of plan area
L′ = water rate, lb/(hr) ft^2

Cooling tower data [15, 19, 46] have been plotted as KaV/L′ vs. L'/G_a and indicate that the tower characteristic KaV/L′ is a function of L'/G_a and not dependent on the value of G_a only, when using high voidage splash deck grid type packings. A few representative tower fill packings are shown in Figures 14-124 A and B, and performance characteristic values are shown in Figure 14-125. Figure 14-126Figure 14-127Figure 14-128Figure 14-129Figure 14-130 and Figure 14-131 illustrate counterflow tower performance. The curves are satisfactory for close estimating, while exact data should be obtained from the manufacturers.

Figure 14-124

(A) Commercial cooling tower fill packing. Reproduced by permission of the American Institute of Chemical Engineers, Kelly, N, W., and Swenson, L. K., *Chemical Engineering Progress*, V. 52, No. 7 © (1956) p. 263; all rights reserved. (B) Representative generic types of fill for cooling towers.

Used by permission of the American Institute of Chemical Engineers, Beyer, A. H., *Chemical Engineering Progress*, V. 89, No. 7 © (1993); all rights reserved.

Figure 14-125

Typical effect of hot water temperature on tower characteristic, KaV/L′ at constant L′, G_a wet bulb temperature and packed height. Note L and G shown in chart are hourly rates.

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Figure 14-126

Percent change in gpm to produce 1°F change in approach.

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Figure 14-127

Effect of half-speed operation of fans.

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Figure 14-128

Decrease from approach for maintaining design water temperature.

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Figure 14-129

Fan efficiency versus specific speed for various blade settings of fans.

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Figure 14-130

Rpm correlation for 16° pitch angle of fan blade.

Used by permission of Whitesell, J., *Chemical Engineering*, Jan. (1955) p. 187, all rights reserved.

Figure 14-131

Diameter correlation for 16° pitch angle for fan blades.

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The fill illustrated in Figures 14-124 A and B is typical of many cooling tower heat transfer evaporative cooling surfaces. The wooden splash type is the oldest in terms of length of usage, while the film types (some fabricated of plastic) have been in service about 40 years [148].

This latter type appears similar to some previously discussed, closely spaced structural packing, but is specifically designed for this application. Beyers [148] recommends film fill as the best choice if the water conditions of Table 14-52 are appropriate. For scaling or plugging water conditions, select splash fill.

Table 14-52 Guidelines for Cooling Tower Recirculating Water (Normal Limits for Using Film Fill)

Reproduced by permission from American Institute of Chemical Engineers, Beyer, A. H., *Chem. Eng. Prog.*, Vol. 89, No. 7 © (1993), p. 42; all rights reserved.

pH—Ideally 6.5-8.0; pH as low as 5.0 is acceptable if galvanized steel is not present
Chlorides—Maximum 750 ppm (as NaCl) for galvanized steel; maximum 1,500 ppm for Type 300 stainless steel; maximum 4,000 ppm for Type 316 stainless steel; silicon bronze is the preferred material if chlorides exceed 4,000 ppm
Calcium—In general, calcium (as $CaCO_3$) below 800 ppm should not result in calcium sulfate scale. In arid climates, however, the critical level may be much lower. For calcium carbonate scaling tendencies, calculate the Langelier Saturation Index or the Ryznar Stability Index
Sulfates—If calcium exceeds 800 ppm, sulfates should be limited to 800 ppm, less in arid climates, to prevent scale. Otherwise, a sulfate level up to 5,000 ppm is acceptable
Silica—Generally, limit silica to 150 ppm as SiO_2 to prevent silica scale

Iron—Limit to 3 ppm. Note that excessive concentrations of iron may stain cooling tower components, but these stains are not the result of any rust or corrosion
Manganese—Limit to 0.1 ppm
Total Dissolved Solids (TDS)—Over 5,000 ppm can adversely affect thermal performance and may be detrimental to wood in the alternately wet/dry areas of the tower
Suspended Solids—Limit to 150 ppm if the solids are abrasive. Avoid film fill if solids are fibrous, greasy, fatty, or tarry
Oil and Grease—Over 10 ppm will cause noticeable thermal performance loss
Nutrients—Nitrates, ammonia, oils, glycols, alcohols, sugars, and phosphates can promote growth of algae and slime. This growth can cause tower problems, particularly with film fill
Ammonia—Limit to 50 ppm if copper alloys are present
Organic Solvents—These can attack plastics and should be avoided
Biological Oxygen Demand (BOD)—Limit BOD to 25 ppm, particularly if suspended solids exceed 25 ppm.
Sulfides—Should be limited to 1 ppm
Langelier Saturation Index—Ideally, maintain between −0.5 and +0.5 A negative LSI indicates corrosion tendencies. A positive LSI indicates $CaCO_3$ scaling tendencies

Enthalpy of Air Operating Line

The enthalpy of air at any point on the operating line is [34]:

(14-147)

The equation for the line at terminal conditions is:

(14-148)

where h_1* = enthalpy of air at any temperature higher than inlet, Btu/lb dry air; note that h_1 is exit air

h_2 = enthalpy of inlet air to tower, equivalent to enthalpy of saturated air at wet bulb temperature, Btu/lb dry air, from *Moist Air Tables*, ASHVE Guide.

t_2' = any water temperature lower than inlet water temperature and higher than inlet air wet bulb temperature, °F

t_1 = inlet water temperature, °F

The effects of wet bulb, approach and range on mechanical draft cooling tower size is indicated in Figure 14-132.

Figure 14-132

Effect of cooling tower performance variables on plan ground area required.

Used by permission of Foster Wheeler Corp., Cooling Tower Dept.

The curves are necessarily the approximate midrange of a spread or band of the magnitude of the respective influences on the ground area. That is, the information is good for guidance over the direction certain changes will take in the final selection. For example, the data are referenced to a 70°F wet bulb, and a 15°F approach, therefore, a change in wet bulb only to 75°F will indicate a tower requiring 90% of the ground area. If the approach changes too, then its correction must also be multiplied against the previous result, and the same handling applies to the wet bulb.

In examining the tower performance, it is not the air temperature that determine capacity, but the heat content or enthalpy of the air. Although the air temperature and wet bulbs at inlet may be different for two different inlet air conditions, it is still possible for the air to have the same enthalpy. Therefore, two different air streams of different properties can produce the same effect on tower performance. The heat content, or enthalpy, of all air with the same wet bulb is the same, therefore it is clear that the wet bulb temperature is important in determining performance.

The data in public literature are limited to performance evaluation of existing cooling towers. It does not allow the design engineer to actually design the height of tower packing or “fill”. Even though the number of transfer units can be estimated (as in Reference 145), these numbers cannot be used in design, because there are essentially no published values for converting the gas mass transfer to square feet of effective contact tower filling, other than the proprietary data of the respective manufacturers [146]. Some data are available for relating the sizes of individual “cells” of respective manufacturers (a cell is a unit size containing x-number of square feet of fill surface distributed throughout y-number of vertical decks in the cell). These are not standard across manufacturers, because the mass transfer of the respective fill varies with specific design of contact surface between the water surface and the airflow.

Mechanical draft towers are normally designed for L/G (liquid/gas rate) of 0.75 to 1.50 [147], and the values of KaV/L vary from 0.50 to 2.50.

Recent studies indicate that the performance of all commonly used commercial high voidage packings can be correlated by [38, 146]:

(14-149)

where L′ = total water flow, lb/hr
N′ = no. of deck levels in tower
t_1 = water temperature at bottom of tower, °F
t_2 = water temperature at top of tower, °F
t_L = water temperature of bulk of water, °F
V = tower volume, ft^3/ft^2 plan area
i_G = enthalpy of air saturated at wet bulb temperature, Btu/lb dry air
i_L = enthalpy of air saturated at bulk water temperature, Btu/lb dry air
K = overall enthalpy transfer coefficient, lb/hr (ft^2 transfer area) (lb water/lb dry air)

This relates the tower characteristic to the number of packing decks in the tower and the L'/G_a ratio. Values of A′ and n′ are given in Table 14-53.

Table 14-53 Values of A′, n′, B, C′, and S_F

Used by permission: The American Institute of Chemical Engineers; Kelly, N. W. and Swenson, L. K, *Chem. Eng. Prog.*, Vol. 52 © (1956) p. 263 [38]; all rights reserved.

				B	C′
Deck type	**A′**	**n′**	**S_F, ft**	**Mult. × 10^{-8}**	**Mult. × 10^{-12}**
A	0.060	0.62	3.00	0.34	0.11
B	0.070	0.62	4.00	0.34	0.11
C	0.092	0.60	3.75	0.40	0.14
D	0.119	0.58	6.00	0.40	0.14

D	0.115	0.58	6.00	0.40	0.14
E	0.110	0.46	4.95	0.60	0.15
F	0.100	0.51	9.13	0.26	0.07
G	0.104	0.57	6.85	0.40	0.10
H	0.127	0.47	3.64	0.75	0.26
I	0.135	0.57	4.50	0.52	0.16
J	0.103	0.54	6.85	0.40	0.10

The simultaneous solution of Equation 14-147 involving the approach and cooling range, and Equation 14-149 involving the number of packing decks (and thereby available surface area) yields the $L'G_a$, which satisfies the specified performance. The accuracy of this combined with the data is within 5%. Equation 14-149 is essentially a straight line on log-log paper, so two points are sufficient to determine its position. The lengthy step of the process is integrating the expression for the several enthalpy conditions involving approach and range that could satisfy the problem.

14.50. Ground Area vs. Height

The economics of forced and induced draft cooling tower operation require a study of fan and water pump horsepower, and usually dictate a fan static pressure requirement that does not exceed 0.75–1.0 in. of water. For atmospheric and natural draft towers, the economics of pumping water are still very important. This means that the ground area must be so selected as to keep the height down, while not dropping the unit rates so low that performance becomes poor. This then, is a balance of ground area versus total deck height. Pritchard [200] presents an estimation curve, indicating that as packed height varies from 12–40 ft, the economics of ground area suggest a G_a of 2,000–1,400 respectively, being slightly less than a straight line function.

14.51. Pressure Losses

The tower pressure losses are: (1) tower packing or fill (70–80% of loss); (2) air inlet if induced draft; (3) mist eliminators at top; (4) air direction change losses and entrance to packing on forced draft units. These losses are a function of air velocity, number and spacing of packing decks, liquid rate and the relation between L′ and G_a.

The pressure drop for a given number and type of packing deck is expressed as [38]:

(14-150)

Values of B, C and Sp are given in Table 14-53. Pressure drop values, ΔP′/N′, per individual deck range from 0.003–0.006 in. water for low L′ and G_a rates to 0.03–0.06 in. water for high L′ (3,500) and G_a (2,000) rates [19]. Values of G_E are taken from Figure 14-133. A typical pressure drop curve is shown in Figure 14-134.

Figure 14-133

Values of equivalent air mas velocity.

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Figure 14-134

Typical effect of liquid loading on air pressure drop showing linear relationship; total deck height is 20 ft.

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Pressure loss through wooden mist eliminators based on 0.0675 lb/ft^3 air varies from 0.01 in. water at G_a = 800 to 0.07 at G_a = 2,000 as almost a straight line function [16]. These losses are based on the face area of the eliminators.

Pressure loss for inlet louvers based on two face velocity heads and 0.075 lb/ft^3 air is given as 0.02 in. water for 400 ft/min face velocity to 0.32 in. water for 1,600 fpm, varying slightly less than a straight line [19].

14.52. Fan Horsepower for Mechanical Draft Tower

(14-151)

where F = actual cfm at fan inlet, ft^3/min
p_s = total static pressure of fan, in. of water

This relation includes a 50% static efficiency of the fan and gear losses, assuming a gear drive [19]. If belt driven the difference will not be great.

For study purposes, the effects of performance as related to fan horsepower may be patterned after Figure 14-135 and Figure 14-136. The conditions for actual air inlet for an induced draft fan must be obtained from Equation 14-147 read from a diagram similar to Figure

14-123.

Figure 14-135

Fan power requirements for components of a typical counterflow induced draft cooling tower.

Used by permission of Groitein, E. E., *Combustion*, Nov. (1957) p. 38.

Figure 14-136

Plot illustrating combination of thermal and fan power characteristics to directly determine flow capacities of a given size.

Used by permission of Groitein, E. E., *Combustion*, Nov. (1957).

Economical tower sizes usually require fan horsepower between 0.05 and 0.08 hp/ft^2 of ground plan area [19], and motors larger than 75 hp are not often used due to inability to operate proper fans and gears in the space required.

14.53. Water Rates and Distribution

Water distribution must give uniform flow over the tower packing. Many towers use a gravity feed system, discharging the water through troughs and ceramic, metal or plastic nozzles. Other systems use pressure nozzles discharging upward, so that water flows upward

before falling back over the packing. This latter method requires more pumping head due to the pressure required at the nozzles. Water rates usually run from 1 to 3.5 gpm/ft^2 of ground plan area.

14.54. Blowdown and Contamination Buildup

As the circulating water evaporates whilst passing through the tower, the evaporated water vapor is pure. This means that any material dissolved in the remaining water increases in concentration. This concentration can aggravate the heat transfer surfaces, and develop corrosive conditions on many mechanical and structural parts of the tower. To limit this buildup, a certain amount of liquid is blown down to expel the concentrated material, and this is then replaced with fresh make-up water. See Figure 14-137 for blowdown.

Figure 14-137

Cooling tower blowdown versus number of concentrations in circulating water; percent of total circulating water to be wasted to maintain various concentrations.

Used by permission of Hensley, J. C. (ed.) Cooling Tower *Fundamentals*, 2nd Ed. (1985), The Marley Cooling Tower Co., a United Dominion Co.

The level to which the contamination can concentrate in the circulating water is [44]:

(14-152)

(14-153)

where C = contaminant level in circulating water; number of concentration ratios

E = rate of evaporation, gpm (if not accurately known, evaporation can be approximated by multiplying total water rate in gpm times the cooling range (°F) times 0.0008).

(14-154)

where gpm_T = total cooling tower water flow rate, gpm (incoming to be cooled by tower), DL = drift loss, water lost from tower system entrained in exhaust air stream, measured as (a) % of circulating water rate, gpm, or (b) more precise [44], an L/G parameter and drift becomes pounds of water per million pounds of exhaust air; for estimating:

(14-155)

CR = cooling range, °F, difference between hot water into tower and cold water from the tower, °F

B = rate of blowdown, gpm. (Because an acceptable level of concentration has usually been predetermined, the operator is more concerned with the amount of blowdown necessary to maintain the concentration, Equation 14-152.)

L/G = ratio of total mass flow of water and dry air in cooling tower, lb/lb

Example: 14-17: Determining Approximate Blowdown for Cooling Tower (Used by permission of Marley Cooling Tower Co., Inc., from Cooling Tower

Fundamentals [144])

Assume that a given cooling tower is designed to reduce the incoming temperature of 10,000 gpm by 25°F (range). Then, assume that the level of chlorides in the make-up water is 250 ppm, and that level should not exceed 750 ppm in the circulating water. Allowable concentration ratio is 750/250 = 3. The approximate evaporation rate would be 10,000 × 25 × 0.0008 = 200 gpm. The approximate drift rate would be 10,000 × 0.0002 = 2 gpm. Applying these values to Equation 14-153, blowdown would be:

Even if the assumed evaporation and drift rates were perfectly accurate, the calculated blowdown rate of 98 gpm might still not be quite enough because of the effects of air-borne contaminants, which are usually incalculable. Once the approximate level of blowdown has been determined, the circulating water quality should be regularly monitored and appropriate adjustments made.

Figure 14-137 is a plot of the percentage of circulating water flow to be wasted in order to maintain various concentrations, based upon the approximate evaporation and drift rates indicated by Equations (14-71) and (14-72), expressed as percentages.

Despite the benefits of blowdown, however, chemical, electrostatic, or electronic treatment of the water is often required to prevent scale formation, corrosion, or biological growth. When treatment is required, or anticipated to be required, the services of a reliable water treatment company should be obtained.

Brooke [149] provides calculation techniques using enthalpy of the air to determine water evaporated, air flow, and blowdown quantities.

14.55. Preliminary Design Estimate of New Tower

Refer to psychrometric chart, Figure 14-138A and Figure 14-138B, for basic considerations in establishing tower conditions.

1. Determine the inlet water temperature to the tower. This is approximately the outlet temperature from the cooling water load.
2. Determine the heat load to be performed by the tower, based on required water inlet and outlet temperatures and flow rates.
3. Establish the wet bulb temperature for the air at the geographical site of the tower. Use weather bureau records if other data are not available. Use caution; do not select too high a value.
4. Prepare a plot of the saturation curve for air-water. Establish the operating line by starting at the point set by the outlet cold water temperature and the enthalpy of air at the wet bulb temperature, and with a slope L'/G_a assumed between 0.9 and 2.7. See Figure 14-123.
5. Graphically integrate, by plotting $l/h'-h$ vs. t, reading $(h'-h)$ from the operating-equilibrium line plot for various values of temperature. See Figure 14-139.

Figure 14-139

Graphical integration to determine number of transfer units.

6. The value of the integral is equal to the number of transfer units, so set it equal to Equation 14-149 and solve for the number of decks needed, N. Select the desired deck from Figures 14-124

A and B and the constants A′ and n from Table 14-53.

7. If the number of decks required is unreasonable from a height standpoint, the procedure must be repeated using a new assumed L'/G_a, or a new approach, or a new wet bulb temperature, or some combination of these.

8. For the assumed L'/G_a and known L′, calculate the required air rate G_a.

Figure 14-138A Psychrometric chart, reference barometric pressure of 29.92 in. Hg.

Used by permission of Westinghouse Electric Co., Sturtevant Div.

Figure 14-138B

Directions for using Figure 14-124 A.

Used by permission of Westinghouse Electric Co., Sturtevant Div.

14.56. Alternate Preliminary Design of New Tower (after References 12 and 19)

1. Follow Steps (1), (2), and (3) of the procedure just outlined.

2. Refer to a plot of KaV/L′ versus L'/G_a as in Figure 14-140AFigure 14-140BFigure 14-140CFigure 14-140DFigure 14-140EFigure 14-140F and Figure 14-140G or in References 15

and 19. This saves the integration step, as this has been performed for a selection of reasonable wet bulb temperatures and temperature ranges. The curve appropriate to fit the design problem must be used.

Figure 14-140A 65°F wet bulb; 30°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance*, The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

Figure 14-140B

70°F wet bulb; 20°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance*, The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

Figure 14-140C

75°F wet bulb; 25°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance*, The Pritchard Corp. (now, Black and

Veatch Pritchard Corp.) (1957).

Figure 14-140D

75°F wet bulb; 40°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance,* The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

Figure 14-140E

80°F wet bulb; 20°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance,* The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

Figure 14-140F

80°F wet bulb; 30°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance*, The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

Figure 14-140G

80°F wet bulb; 40°F range, counterflow coollng tower performance curves.

Used by permission of *Counterflow Cooling Tower Performance*, The Pritchard Corp. (now, Black and Veatch Pritchard Corp.) (1957).

3. Plot Equation 14-149 for two assumed L'/G_a values and an assumed number of checks on the plot of Figure 14-140C or its equivalent. The intersection with the approach curve gives the value of L'/G_a which satisfies the two Equations (14-138) and (14-149).
4. From the known liquid rate L' and the value of L'/G_a assumed, calculate the needed value of the air rate, G_a. This value is converted to CFM at the fan inlet, together with the calculated pressure drop gives the fan horsepower requirements.

14.57. Performance Evaluation of Existing Tower [19]

1. Because the heat load, L', G_a and temperatures are known for an operating tower, its performance as represented by the number of transfer units, or tower characteristics can be determined. Solve Equation 14-149 for Ka V/L′, or use the modified Merkel diagram, Figure 14-141. This is the number of transfer units

operating in the tower. For relative comparison of Ka values see Figure 14-142.

Figure 14-141

Calculation of KaV/L′ factor.

Used by permission of Brooke, M., *Refining Engineer* May (1958) p. c-41; as reproduced from Woods and Betts, *The Engineer* (London), Mar. 17 and 24 (1950).

Figure 14-142

Comparison of cooling efficiency of several packing materials in terms of the coefficient of heat transfer K′a.

Used by permission of Plastics Technical Service, The Dow Chemical Co., Midland Mich. with data added from Fuller, A. L, *et al. Chemical Engineering Progress*, V. 53, No. 10 (1957) p. 501; all rights reserved.

2. If it is desired to evaluate a change in performance on an existing tower, where the required conditions and numbers of decks and kind of packing are known, calculate KaV/L′ for assumed values of L'/G_a.
3. Following Reference 19, plot this on the appropriate curve (good up to altitudes of 3,000 ft) for KaV/L′ vs. L'/G_a for the proper wet bulb, range and at the intersection of the straight line plot with the approach value selected or needed, read the L'/G_a required to meet the performance conditions.

4. Calculate the new G_a, assuming that L' is known. If on the other hand, it is desired to determine just how much cooling can be obtained, then for a fixed air rate, calculate the L' that can be accommodated.

Example 14-18: Wood Packed Cooling Tower with Recirculation, Induced Draft

Perform the preliminary design on a cooling tower to establish its performance and size.

Required gpm = 5,000
Inlet hot water = 110°F
Outlet cold water = 85°F
Wet bulb = 75°F
Recirculation allowance = 3%

Solution

Use alternate design procedure:

1. Range = 110−85 = 25°F

2. Wet bulb

Because recirculation is to be considered, the ambient wet bulb of 75°F must be corrected.

3. Solve Equation 14-149

$KaV/L' = 0.07 + AN\ (L/G_a)^{-n}$
Select Deck "A":
Constants:
$A' = 0.060$
$n = 0.62$
$L'/G_a = 1.00$ assumed
$N' = 30$, assumed number deck levels
$KaV/L' = 0.07 + 0.060\ (30)\ (1.00)^{-0.62}$
$= 1.87$
Second solution of Equation 14-149 to determine line for plot:
$L'/G_a = 2.0$ assumed
All other values remain the same.
$Ka\ V/L' = 0.07 + 0.060\ (30)\ (2.0)^{-0.62}$
$= 0.07 + (0.060)\ (30)\ (0.651)$
$KaV/L' = 1.24$

4. Plot points

KaV/L′ = 1.87 and 1.24 on Figure 14-140C representing
75°F wet bulb, and 25°F range:
For approach = 85° – 75°F = 10°F
L'/G_a at intersection of straight line plot = 1.13
This is L'/G_a required for 75°F wet bulb.
Enthalpy of exit air, $h_1 = h_2 + (L'/G_a)(t_1 - t_2)$
h_2 at 75°F = 38.61 Btu/lb dry air
$h_1 = 38.61 + 1.13 (110 - 85)$
= 66.91 Btu/lb dry air
Recirculation of 3%:
For 3% of air entering recirculated from the exit air, 97% comes from fresh air.
Enthalpy of recirculated air = 66.91
Enthalpy of fresh air = 38.61
Average enthalpy of inlet mixture:
= 0.97 (38.61) + 0.03 (66.91)
= 39.41 Btu/lb dry air
Refer now to "Moist Air" tables or other data of enthalpy vs. temperature:
At enthalpy = 39.41 Btu/lb dry air, read
Corresponding wet bulb temperature = 76°F (close)
New approach for tower design = 85° – 76° = 9° F, instead of the previous 10°F.
Referring back to plot of number of decks on KaV/L′ vs. L'/G_a,
Read at intersection of 9°F approach, $L'/G_a = 1.05$
5. Estimated G_a, for ground plan area
The assumed 30 decks on 9-in. spacing give a packed height of (30–1) (9/12) = 21.8 ft = 21 ft, 10 in. By approximate straight line interpolation given under "Ground Area Vs. Height":

limit values are 2,000 and 1,400

x = 390 lb/hr (ft^2 ground area) incremental air rate

Suggested G_a = 1,400 + 390 = 1,790 lb/hr (ft^2)

Then for L'/G_a = 1.05

L' = (1,790) (1.05) = 1,880 lb/hr (ft^2)

For 5,000 gpm:

The mass flow rate in lb/hr is: = (5,000) (8.33 lb/gal) (60) = 2,500,000 lb/hr

6. Cooling tower cells

Because cells are in modules of 6 ft, try combination of
30 ft × 24 ft = 720 ft^2
Two cells = 1,440 ft^2
Using this area:

This is about as close as can be estimated without manufacturer's data.

7. Pressure drop through packing

$N' = 30$
$B = 0.34 \times 10^{-8}$
$C = 0.11 \times 10^{-12}$
$S_F = 3.00$
$G_a^2 = (1{,}658)^2 = 2{,}750{,}000$
$L' = 1740$
$G_E = 4{,}050$ at $S_F = 3.0$ and $G_a = 1{,}658$ from Figure 14-133
$\rho_G = 0.07125$ lb/ft^3 avg. for tower

This is an acceptable value.
For pressure drop comparisons of some fills, see Figure 14-143.

Figure 14-143

Comparison of pressure drops of several packing materials.

Used by permission of Plastics Technical Service. The Dow Chemical Co., Midland Mich. with data added from Fuller, A. L., *et al. Chemical Engineering Progress*, V. 53, No. 10 (1957) p. 501; all rights reserved.

8. Pressure loss through louvers for induced draft tower. Assume

louvers are along 24-ft dimension. Total louver face area = (24) (2 cells) (2 sides) (6 ft high) = 576 ft^2 ft^2 tower ground area = (24) (30) (2) = 1,440 Air rate, lb/hr = (1,658) (1,440) = 2,380,000

Note the average value of ρ_G = 0.07125 could be used here.

Face velocity through louvers

Louver pressure drop approximation from paragraph "Pressure Losses"

x = 0.17 in. water

Pressure loss = 0.32 – 0.17 = 0.15 in. water

Note that this value is high by 20–40% due to the approximation.

9. Pressure loss through mist eliminators Approximation will give results on high side:

x = 0.0171 in. water

Pressure loss = 0.07 – 0.02 = 0.05 in. water

10. Total estimated static pressure loss

0.42 + 0.15 + 0.05 = 0.62 in. water
This is an acceptable and reasonable value.

11. Estimated fan brake horsepower

Assume gear drive
Air density at outlet = 0.067 lb/ft^3 (close to 95% saturation at hot water temperature, induced draft fan condition).

This would be in at least two fans, one per cell.

Use 60 hp motor each.

Cooling Tower Based on Ka Data

The data on cooling water with air are usually presented in the literature as Ka values. In using this information the units should be checked very carefully.

1. Height of packing

(14-151)

where K'_a = enthalpy coefficient of total heat transfer, Btu/(hr) (ft^3 tower packing volume) (enthalpy difference between air and water)

2. Number of transfer units

(14-146)

3. K′a Data

Figure 14-142 presents data for several packings in water cooling service.

4. Pressure Drop

Data is given in Figure 14-143 for water–air system.

Performance of Atmospheric and Natural Draft Towers

The evaluation of atmospheric and natural draft towers has not been completely presented in as much as detail mechanical draft towers. Some data are available in estimation form, but the evaluation of transfer rates is only adequate for estimating purposes [4]. The design of such towers must be made only after due consideration of this, and ample safety factors should be included. Figure 14-144 presents general information on water loss due to wind on the tower.

Figure 14-144

Atmospheric cooling tower; water loss for various wind velocities.

Used by permission of Plastics Technical Service, The Dow Chemical Co., Midland Mich. with data added from Fuller, A. L, *et al. Chemical Engineering Progress*. V. 53, No. 10 (1957) p. 501; all rights reserved.

Nomenclature

A = Tower crosssection, ft^2, or riser area, ft^2

A′ = Constant in cooling tower performance equation, Equation 14-149, or in mass transfer = L_m/mG_m

A_F = Final column inside net area, ft^2. or $in.^2$

a = Surface area of an orifice, $in.^2$

a = Effective interfacial area for contacting gas and liquid phases,

ft^2/ft^3. Because this is very difficult to evaluate, it is usually retained as a part of the coefficient such as K_Ga or K_La

a = Area of transfer surface per unit of tower volume in water cooling towers, ft^2/ft^3 or, termed contact area

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a = Area of transfer surface per unit of tower volume in water cooling towers, ft^2/ft^3, or, termed contact area.

A_o = Column (tower) net inside operating area, ft^2.

a/ε^3 = Packing factor, F, not equivalent to experimental values reported.

a′ = Wetted packing surface, not necessarily same as effective interfacial surface area for contact, ft^2/ft^3.

a_t = Total specific packing surface area, ft^2/ft^3.

ACFM = Actual air volume, ft^3/min, or = F

B = Rate of blowdown, (cooling tower), gpm; or = constant in pressure drop equation for cooling tower.

BHP = Brake horsepower, hp.

C = Concentration, lb mol/ft^3.

C′ = Constant in pressure drop equation for cooling towers

C_o = Capacity rating, ft/sec, or m/sec (Norton Chart).

C_o = Constant specific to a particular packing, also,.

C_1 = Constant specific to a particular packing.
$C_3 = 2.4 \times 10^{-8}$.
$C_4 = 2.7 \times 10^{-5}$.
C_f = Vapor capacity factor, (vapor velocity) [(vapor density)/(liquid density – vapor density)]$^{0.5}$
C_{lm} = Log mean of the driving force at the top and bottom of tower.
C_p = Correction factor for high gas ratios, see Figure 14-106.
C_s = Capacity factor = C factor = C, ft/sec.
C_{sc} = Efficient capacity, ft/sec, or m/s. (Norton Chart).
CR = Cooling range (cooling tower), °F.
c = Concentration of solution, lb solute/ft^3 solution.
c, c_0, c_1, c_2, c_3 = Correlation constants.
D = Tower diameter, in. or ft as appropriate.
D_l = Density of liquid, lb/ft^3 = 1 or = L
DL = Rate of drift loss (cooling tower), gpm.
D_L = Diffusivity of solute in liquid, ft^2/hr; or = D_l.
D_g = Gaseous diffusion coefficient, ft^2/hr.
D_p = Equivalent spherical diameter of particle as a sphere possessing same surface area as a piece of packing, ft.
D'_p = Nominal packing size, ft.
D_v = Diffusivity of solute in gas, ft^2/hr.
D_v = Density of vapor, lb/ft^3.
d = Diameter of hole, in.; or orifice diameter, in.
d_i = Individual packing piece inside diameter, in.
d_o = Individual packing piece outside diameter, in.
d_p = Equivalent spherical diameter of particle as a sphere possessing the same surface area as a piece of packing, in.
E = Rate of evaporation (cooling tower), gpm.
E = Overall column efficiency, %.

e = Base of natural logarithms.
F = Packing factor; experimentally determined, ft^{-1}.
F = Actual cfm at fan inlet.
F′ = a/ε^3, characteristic of packing size and shape, not packing factor.
FP = Flow parameter, = F-factor, = F.
F_p = Packing factor, empirical, dimensionless.
F_{pd} = Dry-bed packing factor, dimensionless.
F_{rl} = Liquid Froude Number.
F_s = Superficial F-factor, $V_s (\rho_v)^{0.5}$, ft/sec $(lb/ft^3)^{0.5}$.
f_a = Fraction of total packing area, a_t, wetted.
G = Superficial mass gas rate, lb/(sec)(ft^2 tower crosssection) or lb/hr, lb/sec, or lb/hr-ft^2, lb mol/(sec) (ft^2), depending on equation units.
G′ = Superficial mass gas (vapor) rate, lb/hr (ft^2 tower crosssection), or lb/hr.
G′ = Gas rate, lb/sec.
G_a = Air mass velocity, lb dry air/(hr) (ft^2 or plan area).
GA = Design vapor flow rate, lb/hr.
G_E = Equivalent mass air velocity for pressure drop between air and falling water drops, (lb/hr)/ft^2 of plan area, see Figure 14-133.
G_f = Gas loading factor.
G_m = Gas molar velocity, lb mol/(hr) (ft^2).
G_h = Gas mass velocity, lb/hr (ft^2).
g_c, g = Conversion factor, 32.174 lb_m/lb_f (ft/sec^2), acceleration of gravity, 32 ft/sec^2.
H, HTU = Height of a transfer unit, ft or = differential head at orifice, ft. liquid.
H′ = Henry's Law constant, lb mol/(ft^3) (atm), or other units to suit system.

HETP = Height equivalent to a theoretical plate, ft or in. when specified.

H_g = Height of individual gas film transfer unit, ft.

H_G = Same as H_g.

H_l = Height of individual liquid film transfer unit, ft.

H_L = Same as H_l.

H_{OG} = Height of transfer unit, based on overall gas film coefficients, ft or in. as specified.

H_{OL} = Height of transfer unit, based on overall liquid film coefficients, ft.

$(HTU)_{OG}$ = Height of gas phase transfer unit, ft. or in.

h = Enthalpy of the main air stream, Btu/lb.

h = Liquid height over orifice, in. liquid.

h′ = Enthalpy of saturated air film at bulk water temperature, Btu/lb.

h_d = Vapor pressure drop across distributor, in. liquid.

h_l = Liquid hold-up, ft^3/ft^3 packed tower volume.

h_o = Operating hold-up for any liquid, ft^3/ft^3 packing volume.

h_s = Static hold-up for any liquid, ft^3/ft^3 packing volume.

h_t = Total hold-up, ft^3 liquid/ft^3 packing volume.

h_1^* = Enthalpy of air at any temperature higher than inlet, Btu/lb dry air; Note: h_l = exit air.

h_2 = Enthalpy of inlet air to tower, equivalent to enthalpy of saturated air at wet bulb temperature, Btu/lb dry air from *Moist Air Tables, ASHVE Guide*.

h_{ow} = Operating hold-up for water, ft^3/ft^3 packing volume.

h_{sw} = Static hold-up for water, ft^3/ft^3 packing volume.

h_{tw} = Water hold-up, ft^3 liquid/ ft^3 tower volume.

i_L = Enthalpy of air saturated at bulk water temperature, Btu/lb dry air.

i_G = Enthalpy of air saturated at wet bulb temperature, Btu/lb dry air.
j = Constant, Table 14-44.
K = Overall enthalpy transfer coefficient, lb/(hr) (ft^2 transfer area) (lb water/lb dry air; or, mass transfer coefficient, lb water/(hr) (ft^2).
= Enthalpy coefficient of total heat transfer, Btu/(hr)(ft^3 tower packing volume) (enthalpy difference between air and water).
K_G = Gas phase mass transfer coefficient, lb mol/(hr) (ft^2) (atm).
K_Ga = Overall gas mass transfer coefficient, lb mol/(hr) (ft^3) (atm).
K_L = Overall mass transfer coefficient based on liquid phase, lb mol/ (hr) (ft^2) (lb mol/ft^3).
K_La = Overall mass transfer coefficient based on liquid film controlling, lb mol/ (hr) (ft^3) (lb mol/ft^3).
k_G = Gas-phase mass transfer coefficient, lb mol/(hr) (ft^2) (atm).
k_Ga = Individual gas mass transfer coefficient, lb mol/ (hr) (ft^3) (atm).
k_L = Liquid phase mass-transfer coefficient, lb mol/ (hr) (ft^2) (lb mol/ft^3).
k_La = Individual liquid mass transfer coefficient, lb mol/(hr)(ft^3) (lb mol/ft^3).
L = Liquid superficial mass rate, lb/sec (ft^2 tower crosssection), or lb/sec, depends on equation, or lb mol/ft^2, or liquid rate, lb/sec-ft^2, or lb/hr-ft^2.
, L = Liquid (or water rate) superficial mass rate, lb/(hr)(ft^2 tower plan crosssection, when specified), or lb/hr.
L_c = Critical liquid rate for liquid continuous system, at which

the packing becomes completely wetted and tower operation is in the liquid continuous range, lb/(hr) (ft^2).

L_f = Liquid loading factor.

L_h = Liquid mass velocity, lb/(hr) (ft^2).

L_M = Liquid mass rate, lb mol/(hr) (ft^2), or lb mol/hr.

L_{min} = Minimum liquid wetting rate in packed tower, ft^3/(hr) (ft^2 crosssection).

l = Individual packing piece height (or length), in.

m = Slope of equilibrium line, mol fraction.

M = Molecular weight of compound.

M_M = Mean molecular weight of gas, lb/lb mol.

MWR = Minimum wetting rate, liquid, ft^3/(hr) (ft^2 crosssection)/ft^2 packing surface/(ft^3 tower volume).

MWR_G = Minimum wetting rate, gpm/ft^2, (Eq. 14-15).

MWR_T = Minimum wetting rate, gpm/ft^2 , constant, allowing for vertical height consumed by distribution/redistribution equipment, see Table 14-25a.

N = Number of transfer units.

= Number of deck levels in cooling tower.

N_{OG} = Number of transfer units, based on overall gas film coefficients.

N_{OL} = Number of transfer units, based on overall liquid film coefficients.

$(NTU)_{OG}$ = Number of gas transfer units.

n = Moles absorbed or transferred, lb mol/hr; or

n = Number of holes in distributor.

n = Constant in cooling tower performance equation, or constant for Equation 14-145.

P = Average total pressure in tower, atm or absolute units, or = P_{avg}.

= Specific pressure drop through dry bed, in. water /ft packing.

P_{BM} = Mean partial pressure of inert gas in the gas phase, atm.

= Pressure drop, in. water/ft packed height, or in. of liquid; or pressure drop through risers, in. liquid.

= Air pressure loss, in. of water.

= Pressure drop at flood point for all random packings, in. of water/ft. of packing height.

= Dry bed pressure drop, in. water/ft packed height.

= Wet pressure drop, in. liquid/ft. packed height (Nutter).

p = Partial pressure of absolute gas, atm.

p_{lm} = Log mean partial pressure of gas in inlet and exit gas streams, atm.

p_s = Total static pressure of fan, in. of water.

Q = Liquid volumetric flow rate, gpm.

Q_c = Condenser duty, Btu/hr.

Q_r = Reboiler duty, Btu/hr.

Re_g = Reynolds number for gas/vapor.

S = Tray spacing, in. (or ft. when specified).

S = Length of corrugation side (structured packing).

S_F = Vertical free fall of water drops, ft.

Sc_g = Gas phase Schmidt Number.

T = Absolute temperature, $^oR = {}^oF + 460$, ($K = {}^oC + 273\ 15$).

t = Bulk water temperature, ^oF or t_L, if specified.

t_1 = Entering or inlet water temperature, or at the bottom of

tower, ^oF.

t_2 = Outlet water temperature, or at the top of tower, ^oF.

= Any water temperature lower than the inlet water temperature and higher than the inlet air wet bulb temperature, ^oF.

t_L = Water temperature of bulk water, ^oF.

u_g = Superficial velocity, based on crosssection area of empty column, ft/sec.

u_{maxIE} = Calculated or estimated maximum superficial vapor velocity, ft/sec.

u_{max} = Maximum vapor velocity for actual packing, ft/sec.

U_{ge} = Effective velocity of gas.

U_g = Superficial velocity of gas.

U_1 = Superficial velocity of liquid.

V = Tower volume, ft^3/ft^2 ground plan area; or active cooling volume ft^3/ft^2 plan area.

V = Horizontal liquid velocity, ft/sec. (in distributor).

V = Vapor rate, ft^3/sec, or ft/sec.

= Vapor flow, lb mol/hr.

V_L = Liquid rate, gpm

V_g = Superficial gas velocity, ft/sec.

V_A, V_B = Molecular volume of gases, obtained by Kopp's Law of additive volumes, cc/gm mole at normal boiling point, see Table 14-47

V_r = Reboiler vapor rate, lb/hr.

V_s = Superficial vapor velocity, ft/sec (or cross tower crosssection).

V_{SL} = Liquid velocity based on superficial tower area, gpm/ft^2.

V_V = Vapor rate, ft^3/sec.

X = Flow parameter (Norton Co.) = F = FP.

X = Concentration of solute in liquid, lb mol solute/lb mol solute free solvent (or stream).

X∗ = Concentration of solute in liquid, in equilibrium with the gas, lb mol solute/lb mol solvent.

x = Concentration of solute in liquid, mole fraction, or mol fraction of more volatile component in liquid phase.

X_1, X_2 = Curve fit coefficients for C_2, Table 14-35.

X_3, X_4 = Curve fit coefficients for C_3, Table 14-35.

x∗ = Concentration of solute in liquid in equilibrium with gas, mol fraction.

Y = Concentration of solute in gas, lb mol solute/lb mol solute free (solvent) (stream).

Y = Capacity parameter (Norton).

Greek Letters:

α = Relative volatility between components 1 and 2.

= Constants peculiar to packing for dilute and moderate concentrations [74]

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Appendix A. A List of Engineering Process Flow Diagrams and Process Data Sheets

A-1 Process Flow Diagrams Using Visio 2002 Software

Figure 1-12b Process flow diagram (Feed and fuel desulfurization section).

Figure 1-12c Typical process flow diagram for the production of Methyl Tertiary Butyl Ether (MTBE).

Figure 1-14 Piping and instrumentation diagram for Ammonia plant CO_2 removal.

Figure 1-15 Piping and instrumentation diagram: Ammonia synthesis and refrigeration unit (2).

A-2 Process Data Sheets

1. Air cooled heat exchanger process data sheet

2. Centrifugal pump schedule: driver
3. Centrifugal pump schedule: pump

4. Centrifugal pump summary

5. Column schedule

6. Deaerator process data sheet: Deaerated water storage tank
7. Deaerator process data sheet: Deaerator head

8. Drum process data sheet

9. Effluent schedule

10. Equilibrium flash calculation

11. Fabricated equipment schedule

12. Fan/Compressor process duty specification

13. Fractionator calculation summary

14. General services and utilities checklist

15. Hazardous chemical and conditions schedule

16. Heat and mass balances

17. Heat exchanger rating sheet

18. Hydrocarbon dew point calculation

19. Line list schedule

20. Line schedule

21. Line schedule sheet

22. Line summary table

23. Mass balance

24. Mechanical equipment schedule

25. Pipe line list

26. Pipe list

27. Piping process conditions summary

28. Plate heat exchanger data sheet

29. Calculation of pressure drop in fixed catalyst beds

30. Process engineering job analysis summary

31. Pump calculation sheet

32. Pump schedule

33. Relief device philosophy sheet

34. Tank and vessel agitator data sheet

35. Tank process data sheet

36. Tank schedule

37. Tie-in-schedule

38. Tower process data sheet

39. Tray loading summary

40. Trip schedule

41. Utility summary sheet

42. Vessel and tank schedule

43. Vessel and tank summary: driver

44. Vessel schedule

45. Water analysis sheet

Appendix B. Ethics in the Engineering Profession

Ethics in the Engineering Profession

Ethical issues in the engineering profession are becoming more prevalent as companies are competing for the same market with greater demands on engineers or alternatively with increasing demands by the public on the environment (e.g. Three Mile Island, Bhopal, etc.). Appendix B presents ethical, moral, and legal obligations, codes of conduct by engineering institutions. It also presents heuristics (i.e. ways of assisting to discover what is best) and identification strategies that are helpful in solving ethical problems. In the final analysis all heuristics are fallible; however, they provide guidelines toward the solution of a problem as their characteristics are as follows [1]:

- A heuristic does not guarantee a solution.
- It may contradict other heuristics.
- It can reduce the time to solve a problem.
- Its acceptance depends on the immediate context instead of on an absolute standard.

American Society of Mechanical Engineers (ASME)

Ethics

ASME requires ethical practice by each of its members and has adopted the following Code of Ethics of Engineers as referenced in the ASME Constitution, Article C2.1.1.

Code of Ethics for Engineers

The Fundamental Principles

Engineers uphold and advance the integrity, honor and dignity of the engineering profession by:

I. Using their knowledge and skill for the enhancement of human

welfare;

II. Being honest and impartial, and serving with fidelity the public, their employers and clients; and

III. Striving to increase the competence and prestige of the engineering profession.

The Fundamental Canons

1. Engineers shall hold paramount the safety, health and welfare of the public in the performance of their professional duties.
2. Engineers shall perform services only in the areas of their competence.
3. Engineers shall continue their professional development throughout their careers and shall provide opportunities for the professional and ethical development of those engineers under their supervision.
4. Engineers shall act in professional matters for each employer or client as faithful agents or trustees, and shall avoid conflicts of interest or the appearance of conflicts of interest.
5. Engineers shall build their professional reputation on the merit of their services and shall not compete unfairly with others.
6. Engineers shall associate only with reputable persons or organizations.
7. Engineers shall issue public statements only in an objective and truthful manner.
8. Engineers shall consider environmental impact in the performance of their professional duties.

The ASME Criteria for the Interpretation of the Canons

The ASME criteria for interpretation of the Canons are guidelines and represent the objectives toward which members of the engineering profession should strive. They are principles, which an engineer can reference in specific situations. In addition, they provide interpretive

guidance to the ASME Board on Professional Practice and Ethics on the Code of Ethics of Engineers.

1. Engineers shall hold paramount the safety, health and welfare of the public in the performance of their professional duties.

a. Engineers shall recognize that the lives, safety, health and welfare of the general public are dependent upon engineering judgments, decisions and practices incorporated into structures, machines, products, processes and devices.

b. Engineers shall not approve or seal plans and/or specifications that are not of a design safe to the public health and welfare and in conformity with accepted engineering standards.

c. Whenever the Engineers' professional judgments are overruled under circumstances where the safety, health, and welfare of the public are endangered, the Engineers shall inform their clients and/or employers of the possible consequences.

(i) Engineers shall endeavor to provide data such as published standards, test codes, and quality control procedures that will enable the users to understand safe use during life expectancy associated with the designs, products, or systems for which they are responsible.

(ii) Engineers shall conduct reviews of the safety and reliability of the designs, products, or systems for which they are responsible before giving their approval to the plans for the design.

(iii) Whenever Engineers observe conditions, directly related to their employment, which they believe will endanger public safety or health, they shall inform the proper authority of the situation.

d. If Engineers have knowledge of or reason to believe that another person or firm may be in violation of any of the

provisions of these Canons, they shall present such information to the proper authority in writing and shall cooperate with the proper authority in furnishing such further information or assistance as may be required.

2. Engineers shall perform services only in areas of their competence.

a. Engineers shall undertake to perform engineering assignments only when qualified by education and/or experience in the specific technical field of engineering involved.

b. Engineers may accept an assignment requiring education and/or experience outside of their own fields of competence, but their services shall be restricted to other phases of the project in which they are qualified. All other phases of such project shall be performed by qualified associates, consultants, or employees.

3. Engineers shall continue their professional development throughout their careers, and should provide opportunities for the professional and ethical development of those engineers under their supervision.

4. Engineers shall act in professional matters for each employer or client as faithful agents or trustees, and shall avoid conflicts of interest or the appearance of conflicts of interest.

a. Engineers shall avoid all known conflicts of interest with their employers or clients and shall promptly inform their employers or clients of any business association, interests, or circumstances which could influence their judgment or the quality of their services.

b. Engineers shall not undertake any assignments which would knowingly create a potential conflict of interest between themselves and their clients or their employers.

c. Engineers shall not accept compensation, financial or otherwise, from more than one party for services on the same project, or for services pertaining to the same project, unless the circumstances are fully disclosed to, and agreed to, by all interested parties.

d. Engineers shall not solicit or accept financial or other valuable considerations, for specifying products or material or equipment suppliers, without disclosure to their clients or employers.

e. Engineers shall not solicit or accept gratuities, directly or indirectly, from contractors, their agents, or other parties dealing with their clients or employers in connection with work for which they are responsible. Where official public policy or employers' policies tolerate acceptance of modest gratuities or gifts, engineers shall avoid a conflict of interest by complying with appropriate policies and shall avoid the appearance of a conflict of interest.

f. When in public service as members, advisors, or employees of a governmental body or department, Engineers shall not participate in considerations or actions with respect to services provided by them or their organization(s) in private or product engineering practice.

g. Engineers shall not solicit an engineering contract from a governmental body or other entity on which a principal, officer, or employee of their organization serves as a member without disclosing their relationship and removing themselves from any activity of the body which concerns their organization.

h. Engineers working on codes, standards or governmental sanctioned rules and specifications shall exercise careful judgment in their determinations to ensure a balanced viewpoint, and avoid a conflict of interest.

i. When, as a result of their studies, Engineers believe a

project(s) will not be successful, they shall so advise their employer or client.

j. Engineers shall treat information coming to them in the course of their assignments as confidential, and shall not use such information as a means of making personal profit if such action is adverse to the interests of their clients, their employers or the public.

(i) They will not disclose confidential information concerning the business affairs or technical processes of any present or former employer or client or bidder under evaluation, without his consent, unless required by law or court order.

(ii) They shall not reveal confidential information or finding of any commission or board of which they are members unless required by law or court order.

(iii) Designs supplied to Engineers by clients shall not be duplicated by the Engineers for others without the express permission of the client(s).

k. Engineers shall act with fairness and justice to all parties when administering a construction (or other) contract.

l. Before undertaking work for others in which Engineers may make improvements, plans, designs, inventions, or other records which may justify seeking copyrights, patents, or proprietary rights, Engineers shall enter into positive agreements regarding the rights of respective parties.

m. Engineers shall admit their own errors when proven wrong and refrain from distorting or altering the facts to justify their mistakes or decisions.

n. Engineers shall not accept professional employment or assignments outside of their regular work without the knowledge of their employers.

o. Engineers shall not attempt to attract an employee from

other employers or from the marketplace by false or misleading representations.

5. Engineers shall build their professional reputation on the merit of their services and shall not compete unfairly with others.

a. Engineers shall negotiate contracts for professional services on the basis of demonstrated competence and qualifications for the type of professional service required.

b. Engineers shall not request, propose, or accept professional commissions on a contingent basis if, under the circumstances, their professional judgments may be compromised.

c. Engineers shall not falsify or permit misrepresentation of their, or their associates, academic or professional qualification. They shall not misrepresent or exaggerate their degrees or responsibility in or for the subject matter of prior assignments. Brochures or other presentations used to solicit personal employment shall not misrepresent pertinent facts concerning employers, employees, associates, joint ventures, or their accomplishments.

d. Engineers shall prepare articles for the lay or technical press which are only factual.

(i) Technical Communications for publication (theses, articles, papers, reports, etc) which are based on research involving more than one individual (including students and supervising faculty, industrial supervisor/researcher or other co-workers) must recognize all significant contributors. Co authors listed on proposed and accepted publications should have entered the joint authorship arrangement by mutual consent prior to submittal of the document for publication and should have received written permission to use any unpublished work of others

which serves as the major basis or key component of the publication.

(ii) Technical Communications should adhere to clearly defined and appropriately disseminated guidelines on authorship. These guidelines should be promulgated and publicized in corporate, university or other employer policies and should take cognizance of professional technical society recommendations on ethical practice.

(iii) Plagiarism, the act of substantially using another's ideas or written materials without due credit, is unethical.

e. Engineers shall not maliciously or falsely, directly or indirectly, injure the professional reputation, prospects, practice or employment of another engineer, nor shall they indiscriminately criticize another's work.

f. Engineers shall not use equipment, supplies, laboratory or office facilities of their employers to carry on outside private practice without consent.

6. Engineers shall associate only with reputable persons or organizations.

a. Engineers shall not knowingly associate with or permit the use of their names or firm names in business ventures by any person or firm which they know, or have reason to believe, are engaging in business or professional practices of a fraudulent or dishonest nature.

b. Engineers shall not use association with nonengineers, corporations, or partnerships to disguise unethical acts.

7. Engineers shall issue public statements only in an objective and truthful manner.

a. Engineers shall endeavor to extend public knowledge, and to prevent misunderstandings of the achievements of engineering.

b. Engineers shall be completely objective and truthful in all professional reports, statements or testimony. They shall include all relevant and pertinent information in such reports, statements or testimony.

c. Engineers, when serving as expert or technical witnesses before any court, commission, or other tribunal, shall express an engineering opinion only when it is founded on their adequate knowledge of the facts in issue, their background of technical competence in the subject matter, and their belief in the accuracy and propriety of their testimony.

d. Engineers shall issue no statements, criticisms, or arguments on engineering matters which are inspired or paid for by an interested party, or parties, unless they preface their comments by identifying themselves, by disclosing the identities of the party or parties on whose behalf they are speaking, and by revealing the existence of any financial interest they may have in matters under discussion.

e. Engineers shall be truthful in explaining their work and merit, and shall avoid any act tending to promote their own interest at the expense of the integrity and honor of the profession or another individual.

8. Engineers shall consider environmental impact in the performance of their professional duties.

a. Engineers shall concern themselves with the impact of their plans and designs on the environment. When the impact is a clear threat to health or safety of the public, then the guidelines for the Canon revert to those of Canon 1.

9. Engineers accepting membership in The American Society of Mechanical Engineers by this action agree to abide by this Society Policy on Ethics and procedures for its implementation.

Responsibility: Council on Member Affairs/Board on Professional Practice and Ethics

Adopted:	March 7, 1976
Revised:	December 9, 1976
	December 7, 1979
	November 19, 1982
	November 19, 1982
	June 15, 1984
	(editorial changes 7/84)
	June 16, 1988
	September 12, 1991
	September 11, 1994
	June 10, 1998
	September 21, 2002

American Institute of Chemical Engineers (AIChE)

Code of Ethics (Revised January 17, 2003)

Members of the American Institute of Chemical Engineers shall uphold and advance the integrity, honor and dignity of the Engineering profession by: being honest and impartial and serving with fidelity their employers, their clients, and the public; striving to increase the competence and prestige of the engineering profession; and using their knowledge and skill for the enhancement of human welfare. To achieve these goals, members shall

1. Hold paramount the safety, health and welfare of the public and protect the environment in performance of their professional

duties.

2. Formally advise their employers or clients (and consider further disclosure, if warranted) if they perceive that a consequence of their duties will adversely affect the present or future health or safety of their colleagues or the public.

3. Accept responsibility for their actions, seek and heed critical review of their work and offer objective criticism of the work of others.

4. Issue statements or present information only in an objective and truthful manner.

5. Act in professional matters for each employer or client as faithful agents or trustees, avoiding conflicts of interest and never breaching confidentiality.

6. Treat fairly and respectfully all colleagues and co-workers, recognizing their unique contributions and capabilities.

7. Perform professional services only in areas of their competence.

8. Build their professional development throughout their careers, and provide opportunities for the professional development of those under their supervision.

9. Never tolerate harassment.

Conduct themselves in a fair, honorable and respectful manner.

National Society of Professional Engineers

Code of Ethics for Engineers

Preamble

Engineering is an important and learned profession. As members of this profession, engineers are expected to exhibit the highest standards of honesty and integrity. Engineering has a direct and vital impact on the quality of life for all people. Accordingly, the services provided by engineers require honesty, impartiality, fairness and equity, and must be dedicated to the protection of the public health, safety, and welfare.

Engineers must perform under a standard of professional behavior that requires adherence to the highest principles of ethical conduct.

I. Fundamental Canons

Engineers, in the fulfillment of their professional duties, shall:

1. Hold paramount the safety, health and welfare of the public.
2. Perform services only in areas of their competence.
3. Issue public statements only in an objective and truthful manner.
4. Act for each employer or client as faithful agents or trustees.
5. Avoid deceptive acts.
6. Conduct themselves honorably, responsibly, ethically, and lawfully so as to enhance the honor, reputation, and usefulness of the profession.

II. Rules of Practice

1. Engineers shall hold paramount the safety, health, and welfare of the public.
    a. If engineers' judgment is overruled under circumstances that endanger life or property, they shall notify their employer or client and such other authority as may be appropriate.
    b. Engineers shall approve only those engineering documents that are in conformity with applicable standards.
    c. Engineers shall not reveal facts, data or information without the prior consent of the client or employer except as authorized or required by law or this Code.
    d. Engineers shall not permit the use of their names or associate in business ventures with any person or firm that they believe is engaged in fraudulent or dishonest enterprise.

e. Engineers shall not aid or abet the unlawful practice of engineering by a person or firm.

f. Engineers having knowledge of any alleged violation of this Code shall report thereon to appropriate professional bodies and, when relevant, also to public authorities, and cooperate with the proper authorities in furnishing such information or assistance as may be required.

2. Engineers shall perform services only in the areas of their competence.

a. Engineers shall undertake assignments only when qualified by education or experience in the specific technical fields involved.

b. Engineers shall not affix their signatures to any plans or documents dealing with subject matter in which they lack competence, nor to any plan or document not prepared under their direction and control.

c. Engineers may accept assignments and assume responsibility for coordination of an entire project, provided that each technical segment is signed and sealed only by the qualified engineers who prepared the segment.

3. Engineers shall issue public statements only in an objective and truthful manner.

a. Engineers shall be objective and truthful in professional reports, statements, or testimony. They shall include all relevant and pertinent information in such reports, statements, or testimony, which should bear the date indicating when it was current.

b. Engineers may express publicly technical opinions that are founded upon knowledge of the facts and

competence in the subject matter.

c. Engineers shall issue no statements, criticisms, or arguments on technical matters that are inspired or paid for by interested parties, unless they have prefaced their comments by explicitly identifying the interested parties on whose behalf they are speaking, and by revealing the existence of any interest the engineers may have in the matters.

4. Engineers shall act for each employer or client as faithful agents or trustees.

a. Engineers shall disclose all known or potential conflicts of interest that could influence or appear to influence their judgment or the quality of their services.

b. Engineers shall not accept compensation, financial or otherwise, from more than one party for services on the same project, or for services pertaining to the same project, unless the circumstances are fully disclosed and agreed to by all interested parties.

c. Engineers shall not solicit or accept financial or other valuable consideration, directly or indirectly, from outside agents in connection with the work for which they are responsible.

d. Engineers in public service as members, advisors, or employees of a governmental or quasi-governmental body or department shall not participate n decisions with respect to services solicited or provided by them or their organizations in private or public engineering practice.

e. Engineers shall not solicit or accept a contract from a governmental body on which a principal or officer of their organization serves as a member.

5. Engineers shall avoid deceptive acts.

a. Engineers shall not falsify their qualifications or permit misrepresentation of their or their associates' qualification. They shall not misrepresent or exaggerate their responsibility in or for the subject matter of prior assignments. Brochures or other presentations incident to the solicitation of employment shall not misrepresent pertinent facts concerning employers, employees, associates, joint ventures, or past accomplishments.

b. Engineers shall not offer, give, solicit, or receive, either directly or indirectly, any contribution to influence the award of a contract by public authority, or which may be reasonably construed by the public as having the effect or intent of influencing the awarding of a contract. They shall not offer any gift or other valuable consideration in order to secure work. They shall not pay a commission, percentage, or brokerage fee in order to secure work, except to a bona fide employee or bona fide established commercial or marketing agencies retained by them.

III. Professional Obligations

1. Engineers shall be guided in all their relations by the highest standards of honesty and integrity.

a. Engineers shall acknowledge their errors and shall not distort or alter the facts.

b. Engineers shall advise their clients or employers when they believe a project will not be successful.

c. Engineers shall not accept outside employment to the detriment of their regular work or interest. Before accepting any outside engineering employment, they will notify their employers.

d. Engineers shall not attempt to attract an engineer from another employer by false or misleading pretenses.

e. Engineers shall not promote their own interest at the expense of the dignity and integrity of the profession.

2. Engineers shall at all times strive to serve the public interest.

a. Engineers shall seek opportunities to participate in civic affairs; career guidance for youths; and work for the advancement of the safety, health, and well-being of their community.

b. Engineers shall not complete, sign, or seal plans and/or specifications that are not in conformity with applicable engineering standards. If the client or employer insists on such unprofessional conduct, they shall notify the proper authorities and withdraw from further service on the project.

c. Engineers shall endeavor to extend public knowledge and appreciation of engineering and its achievements.

3. Engineers shall avoid all conduct or practice that deceives the public.

a. Engineers shall avoid the use of statements containing a material misrepresentation of fact or omitting a material fact.

b. Consistent with the foregoing, engineers may advertise for recruitment of personnel.

c. Consistent with the foregoing, engineers may prepare articles for the lay or technical press, but such articles shall not imply credit to the author for work performed by others.

4. Engineers shall not disclose, without consent, confidential information concerning the business affairs or technical processes of any present or former client or employer, or public body on which they serve.

a. Engineers shall not, without the consent of all interested parties, promote or arrange for new employment or practice in connection with a specific project for which the engineer has gained particular and specialized knowledge.

b. Engineers shall not without the consent of all interested parties, participate in or represent an adversary interest in connection with a specific project or proceeding in which the engineer has gained particular specialized knowledge on behalf of a former client or employer.

5. Engineers shall not be influenced in their professional duties by conflicting interests.

a. Engineers shall not accept financial or other considerations, including free engineering designs, from material or equipment suppliers for specifying their product.

b. Engineers shall not accept commissions or allowances, directly or indirectly, from contractors or other parties dealing with clients or employers of the engineer in connection with work for which the engineer is responsible.

6. Engineers shall not attempt to obtain employment or advancement or professional engagements by untruthfully criticizing other engineers, or by other improper or questionable methods.

a. Engineers in salaried positions shall accept part-time

engineering work only to the extent consistent with policies of the employer and in accordance with ethical considerations.

b. Engineers shall not, without consent, use equipment, supplies, laboratory, or office facilities of an employer to carry on outside private practice.

7. Engineers shall not attempt to injure, maliciously or falsely, directly or indirectly, the professional reputation, prospects, practice, or employment of other engineers. Engineers who believe others are guilty of unethical or illegal practice shall present such information to the proper authority for action.

a. Engineers in private practice shall not review the work of another engineer for the same client, except with the knowledge of such engineer, or unless the connection of such engineer with the work has been terminated.

b. Engineers in government, industrial, or educational employ are entitled to review and evaluate the work of other engineers when so required by their employment duties.

c. Engineers in sales or industrial employ are entitled to make engineering comparisons of represented products with products of other suppliers.

8. Engineers shall accept personal responsibility for their professional activities, provided, however, that engineers may seek indemnification for services arising out of their practice for other than gross negligence, where the engineer’s interests cannot otherwise be protected.

a. Engineers shall conform with state registration laws in the practice of engineering.

b. Engineers shall not use association with nonengineer, a corporation, or partnership as a “cloak” for unethical acts.

9. Engineers shall give credit for engineering work to those to whom credit is due, and will recognize the proprietary interests of others.

a. Engineers shall, whenever possible, name the person or persons who may be individually responsible for designs, inventions, writings, or other accomplishments.

b. Engineers using designs supplied by a client recognize that the designs remain the property of the client and may not be duplicated by the engineer for others without express permission.

c. Engineers, before undertaking work for others in connection with which the engineer may make improvements, plans, designs, inventions, or other records that may justify copyrights or patents, should enter into a positive agreement regarding ownership.

d. Engineers’ design, data, records, and notes referring exclusively to an employer’s work are the employer’s property. The employer should indemnify the engineer for use of the information for any purpose other than the original purpose.

e. Engineers shall continue their professional development throughout their careers and should keep current in their specialty fields by engaging in professional practice, participating in continuing education course, reading in the technical literature, and attending professional meetings and seminars.

As revised January 2003

"By order of the United States District Court for the District of Columbia, former Section 11 (c) of the NSPE Code of Ethics prohibiting competitive bidding, and all policy statements, opinions, rulings or other guidelines interpreting its scope, have been rescinded as unlawfully interfering with the legal right of engineers, protected under the antitrust laws, to provide price information to prospective clients; accordingly, nothing contained in the NSPE Code of Ethics, policy statements, opinions, rulings or other guidelines prohibits the submission of price quotations or competitive bids for engineering services at any time or in any amount."

Statement by NSPE Executive Committee

In order to correct misunderstandings which have been indicated in some instances since the issuances of the Supreme Court decision and the entry of the Final Judgment, it is noted that in its decision of April 25, 1978, the Supreme Court of the United States declared: "The Sherman Act does not require competitive bidding."

It is further noted that as made clear in the Supreme Court decision:

1. Engineers and firms may individually refuse to bid for engineering services.
2. Clients are not required to seek bids for engineering services.
3. Federal, state, and local laws governing procedures to procure engineering services are not affected, and remain in full force and effect.
4. State societies and local chapters are free to actively and aggressively seek legislation for professional selection and negotiation procedures by public agencies.
5. State registration board rules of professional conduct, including rules prohibiting competitive bidding for engineering

services, are not affected and remain in full force and effect. State registration boards with authority to adopt rules of professional conduct may adopt rules governing procedures to obtain engineering services.

6. As noted by the Supreme Court, "nothing in the judgment prevents NSPE and its members for attempting to influence governmental action…"

Institution of Chemical Engineers (IChemE)

Rules of Professional Conduct

Issue III: 7th December 2001

The Rules and Regulations embodied in this document are made and published as required by By-law 13 of the By-laws of The Institution of Chemical Engineers. They must be read and interpreted in accordance with By-law 1 of that Institution.

Rules of Professional Conduct

1. A code of professional conduct and related Rules designed to cover broad ethical principles must necessarily be drawn up in general terms. These Rules, published by the Council, indicate the manner in which members are required by the Council to conduct themselves in most situations. This conduct must be of a high standard: members must discharge their duties competently, with reasonable skill, and diligence, and shall comply with the standards of behavior, integrity, competence and professional judgement which other members might reasonably expect of them.

For situations not specifically encompassed by these Rules, members are required to order their conduct in accordance with the principle that, in any conflict between a member's personal interests and the fair and proper interests of the community, the member's duty to the community should prevail.

In order to fulfil their duty under these Rules, members shall give due attention to any Guidance Notes that may be published from time to time by the Council in accordance with Rule 2 below.

2. The Council may from time to time publish Guidance Notes to members containing advice as to specific conduct which is to be regarded as proper or improper as the case may be.

3. Members when discharging their professional duties shall act with integrity, in the public interest, and to exercise all reasonable professional skill and care to:

a. Prevent avoidable danger to health or safety.

b. Prevent avoidable adverse impact on the environment.

c.

(i) Maintain their competence.

(ii) Undertake only professional tasks for which they are competent.

(iii) Disclose relevant limitations of competence.

d.

(i) Accept appropriate responsibility for work carried out under their supervision.

(ii) Treat subordinates fairly and without bias.

(iii) Encourage others to advance their learning and competence.

e.

(i) Avoid where possible real or perceived conflict of interests.

(ii) Advise affected parties when such conflicts arise.

f. Observe the proper duties of confidentiality owed to appropriate parties.

g. Reject bribery and any other corrupt practices.

h. Assess relevant risks and liability, and, if appropriate,

hold professional indemnity insurance.

i. Notify the Institution if convicted of a criminal offence or upon becoming brankrupt or disqualified as a Company Director.

j. Notify the Institution of any significant violation of the Institution's Code of Conduct by another member.

k. Be mindful at all times of the dignity of the profession in their personal conduct.

4. Members may be deemed guilty of improper professional conduct if:

a. in the opinion of the Council their conduct fails to conform with the By-laws of the Institution (including the Code of Conduct contained in By-law 12(b)) or of these Rules or with any advice published thereunder; or

b. they are convicted by a competent tribunal of any criminal offence, or have in any other proceedings to which they are party been found by a Court or other competent tribunal guilty of conduct which (in either such case) in the opinion of the Council renders them unfit to be members; or

c. they are, in the opinion of the Council, precluded from performing their professional duties, in a manner which is consistent with the standards of their profession, as a result of having been adjudicated bankrupt or of making a composition with their creditors; or

d. in the opinion of the Council their conduct in other respects renders them unfit to be members of the Institution or is injurious to the Institution.

The following are heuristics that can be used effectively in engineering ethics [2]:

1. Use a traditional problem-solving strategy for solving ethical

problems.
2. Consider the possibility that inexperienced people can be right.
3. Debrief people fully before assuming facts about their actions.
4. Consider what will happen if a specific decision is based on a false assumption.
5. Be honest and open. This is especially important when dealing with those who are predisposed to distrust you.
6. Be concerned about the welfare of your company.
7. Be concerned about the welfare of your employees.
8. Do not let people make ethical decisions for you.
9. Acquire all the information you can about the situation. The problem may be more serious than it first appears, or there may be no real problem at all.
10. Acknowledge the concerns of others, whether you share their concerns or not.
11. Remember that one is only as ethical as one can afford to be. One can enhance one's ability to afford to do the right thing by knowing the consequences and by balancing the responsibilities that one accepts with the rest of one's life.

Woods proposed the McMaster five-step strategy [3] in analyzing standard problem solving strategy [4] as follows:

Define. Was the problem well defined? Was the real problem that now things were more uncertain than before? There are other possible definitions of the problem, but, if one defines the problem as how to deal with obviously flawed data, one may miss the entire point. The problem would be thereby unnecessarily overconstrained. What is the real problem? If the characters had defined the problem better, would they reach a different conclusion? Would it have been a better conclusion?

Explore. Explore the “real problem” by creating a rich perspective of the problem. During the Explore stage, see it from many different view points. Be willing to spend at least half the total available time defining the problem. Ask many “what if?” questions. Identify the real problem, by asking a series of “Why?” questions to generalize the situation and to see the problem in the context of a “system.”

Plan. Use the criteria to select a sequence of diagnostic actions. Sometimes several actions can be combined. However, usually it is best to wait for the results from the first action before we recycle back to the Explore stage and relook our hypotheses. The criteria in selecting an action include the following: Will the action provide background information to ensure the problem is understood in context or is the action to test a hypothesis? Will the action produce results that give the accuracy needed? Is the action simple? inexpensive? safe?

Do It. Carry out the first action in the plan. Check. Monitor.

Look back. Compare the results obtained with the hypotheses.

Look back at the process used. Self-assess. Return to previous stages of *Engage*, *Explore*, *Pla*n and continue.

Blanchard and Peale [5] proposed a set of guidelines that helps to sort out the issues on ethical solution and to consider the following checklist of questions:

1. Is it legal? Will I be violating either civil law or company policy?
2. Is it balanced? Is it fair to all concerned in the short term as well as the long term? Does it promote win-win relationships?
3. How will it make me feel about myself? Will it make me proud? Would I feel good if my decisions were published in the newspaper? Would I feel good if my family knew about it?

If the answer to the first question could be interpreted as “NO, it is not legal,” then there is no need to proceed to the second and third Ethics Check list questions. However, if the solution is indeed legal and does not violate the company policy, then the second question raises the flag that a decision that greatly benefits one individual or company will eventually come back to haunt that individual or company. The last question is directed to prompt our sense of fairness and to ensure that our self-esteem is not diminished through an unethical decision.

In addition to the checklist, Blanchard and Peale discussed the five P’s that can be considered in analyzing the solution: Purpose, Pride, Patience, Persistence, and Perspective.

The Five P’s

Purpose: What is the objective for which you are striving? Are you comfortable with that as your purpose? Does your purpose hold up when you look at yourself in the mirror?

Pride: Can you take pride in the solution you have developed? Is there any false pride or self-doubt involved?

Patience: Have you taken the time to think through all the ramifications of your solution?

Figure B-1.
The interrelationships among the five P’s [4].

Persistence: Are you sticking to your guns and not being dissuaded by other demands? Have you given up too soon on finding a solution that

is fair and balanced to all concerned?

Perspective: Have you taken the time to focus inside yourself to be sure everything fits with your ideals and beliefs? How does the solution fit into the "Big Picture?"

Figure B-1 shows the interrelationships among the five P's. Perspective is the fifth P, the hub around which the other P's rotate. Part of Perspective is the inner guidance that is awakened from the other P's that helps us see things more clearly [4].

The various codes of conduct in this appendix should enable the engineers to make the correct decisions which can be identified by the use of a heuristic. The correct decision is one that is [2]

1. Consistent with the engineer's moral principles.
2. Consistent with the general accepted codes of engineering conduct.
3. Consistent with obligations that the engineer has accepted.
4. Consistent with the law.
5. Consistent with the applicable code of ethics.

Although it is possible that an individual decision may not be acceptable to another individual, ultimately the correct decision is one that individual can live with. The ability to make one's own ethical decisions is known as moral autonomy.

References

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2. R. Turton, R.C. Bailie, W.B. Whiting, J.A. Shaeiwitz, *Analysis, Synthesis and Design of Chemical Processes*. 2nd ed (2003) Prentice Hall Int, NJ .
3. D.R. Woods, *A Strategy for Problem Solving*. 3rd ed (1985) Department of Chemical Engineering, McMaster University, Hamilton, Ontario .

4. H.S. Fogler, S.E. LeBlanc, *Strategies for Creative Problem Solving*. (1995) Prentice Hall, Englewood Cliffs, NJ .
5. K. Blanchard, N.V. Peale, *The Power of Ethical Management*. (1988) Ballantine Books, Fawcett Crest, New York .

Online Ethics Center for Engineering and Science

http://www.onlineethics.org
http://ethics.tamu.edu
http://www.niee.org
http://www.nspe.org

Appendix C. Physical Properties of Liquids and Gases

Tables of Physical Properties of Liquids and Gases

C-1	Density of Liquids
C-2	Viscosity of Gas
C-3	Viscosity of Liquids
C-4	Heat Capacity of Gas
C-5	Heat Capacity of Liquid
C-6	Thermal Conductivity of Gas
C-7	Thermal Conductivity of Liquids and Solids
C-8	Surface Tension of Organic Liquids
C-9	Vapor Pressure
C-10	Enthalpy of Vaporization
C-11	Enthalpy of Formation
C-12	Gibbs Energy of Formation
C-13	Solubility in Salt Water
C-14	Solubility of Organic Compounds in Water as a Function of Temperature

C-15	Henry's Law Constant for Gases in Water as a Function of Temperature
C-16	Solubility of Selected Gases in Water as a Function of Temperature
C-17	Solubility of Sulfur Compounds in Water as a Function of Boiling Point for Mercaptans and Aromatics
C-18	Solubility of Naphthenes in Water
C-19	Solubility of Nitrogen Compounds in Water
C-20	Henry's Law Constant for Nitrogen Compounds in Water
C-21	Coefficient of Thermal Expansion of Liquids
C-22	Adsorption Capacity of Activated Carbon

Table C-1. Density of Liquids

Table C-2. Viscosity of Gas

Table C-3. Viscosity of Liquids

Table C-4. Heat Capacity of Gas

Table C-5. Heat Capacity of Liquids

Table C-6. Thermal Conductivity of Gas

Table C-7a.
Thermal Conductivity of Liquids

Table C-8.
Surface Tension of Organic Liquids

Table C-9.
Vapor Pressure

Table C-10.
Enthalpy of Vaporization

Table C-11. Enthalpy of Formation

Table C-12. Gibbs Energy of Formation

Table C-13. Solubility of Salt Water

Table C-14. Solubility of Organic Compounds in Water as a Function of Temperature

Table C-15. Henry's Law Constant for Gases in Water as Function of Temperature

Table C-16. Solubility of Selected Gases in Water as a Function of Temperature

Table C-17. Solubility of Sulfur Compounds in Water as a Function of Boiling Point for Mercaptans and

Aromatics

Table C-18. Solubility of Napthenes in Water

Table C-19. Solubility of Nitrogen Compounds in Water

Table C-20. Henry's Law Constant for Nitrogen Compounds in Water

Table C-21. Coefficient of Thermal Expansion of Liquids

Table C-22. Adsorption Capacity of Activated Carbon

Further Reading

1. Yaws, C.L., *Chemical Properties Handbook: Physical, Thermodynamic, Environmental, Transport, Safety, and Health Related Properties for Organic and Inorganic Chemicals*, McGraw-Hill, 1995.
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3. Yaws, C.L., J.R. Hopper, Sunil R. Mishra, and R.W. Pike, "Solubility and Henry's Law Constants for Amines in Water", *Chem. Eng.*, Aug 2001, p. 84.
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Appendix D. Alphabetical Conversion Factors

D-1. Alphabetical Conversion Factors

A

Acre	×10	= Square chain (Gunters)
Acre	×160	= Rods
Acre	$\times 1\times 10^{5}$	= Square links (Gunters)
Acre	×0.4047	= Hectare or square hectometer
Acres	×43,560	= Square feet
Acres	×4,047	= Square meters
Acres	$\times 1.562\times 10^{-3}$	= Square miles
Acres	×4,840	= Square yards
Acre-feet	×43,560	= Cubic feet
Acre-feet	$\times 3.259\times 10^{5}$	= Gallons
Amperes/square centimeters	×6.452	= Amperes/square inch
Amperes/square centimeters	$\times 10^{4}$	= Amperes/square meter
Amperes/square inch	×0.1550	= Amperes/square centimeters
Amperes/square inches	×1,550	= Amperes/square meter
Amperes/square meter	$\times 10^{-4}$	= Amperes/square centimeter
Amperes/square meter	$\times 6.452\times 10^{-4}$	= Amperes/square inch
Ares	×0.02471	= Acre (USA)
Ares	×119.60	= Square yards
Ares	×100	= Square meters
Atmospheres	×14.7	= Pounds/square inch
Atmospheres	×1.058	= Tonnes/square foot
Atmospheres	×29.92	= Inches of mercury at 0°C
Atmospheres	×76	= Centimeters of mercury
Atmospheres	×33.90	= Feet of water at 4°C

Atmospheres	×1.0333	= Kilograms/square centimeter
Atmospheres	×10,332	= Kilograms/square meter
Atmospheres	×1,013.2	= Millibar
Atmospheres	×760	= Millimeters of mercury
Atmospheres	×10.332	= Meters of water at 4°C

B

Barrels (USA, dry)	×7,056	= Cubic inches
Barrels (USA, dry)	×105	= Quarts (dry)
Barrels (USA, liquid)	×31.5	= Gallons
Barrels (oil)	×42	= Gallons (oil)
Barrels (oil)	×0.159	= Cubic meters
Barrels (oil)	×159	= Liters
Barrels/day	$\times 6.6245\times 10^{-3}$	= Cubic meters/hour
Barrels (oil)	×5.6154	= Cubic feet
Barrels/day	$\times 29.167\times 10^{-3}$	= Gallons per minute
Bars	×0.9869	= Atmospheres
Bars	$\times 10^{6}$	= Dynes/square centimeter
Bars	$\times 1.020\times 10^{4}$	= Kilograms/square meter
Bars	×2,089	= Pounds/square foot
Bars	×14.50	= Pounds/square inch
Board feet×144 sq.in.	×1in.	= Cubic inches
Board feet	×0.0833	= Cubic feet
Btu	$\times 1.055\times 10^{10}$	= Ergs
Btu	×778.3	= Foot-pounds
Btu	×252	= Gram-calories
Btu	$\times 3.931\times 10^{-4}$	= Horsepower-hours
Btu	×1,054.8	= Joules

Btu	×0.252	= Kilogram-calories
Btu	×107.5	= Kilogram-meters
Btu	$\times 2.928\times 10^{-4}$	= Kilowatt-hours
Btu/hour	×0.2162	= Foot-pounds/second
Btu/hour	×0.070	= Gram-calorie/second
Btu/hour	$\times 3.929\times 10^{-4}$	= Horsepower-hours (British)
Btu/hour	×0.2931	= Watts
Btu/hour foot °F	×1.4882	= Kilocalorie/meter hour °C
Btu/hour foot^2	×2.7125	= Kilocalorie/meter^2 hour
Btu/pound	×0.5556	= Kilocalorie/kilogram
Btu/pound °F	×1	= Kilocalorie/kilogram °C
Btu inches/hour foot^2 °F	×0.12402	= Kilocalorie/meter hour °C
Btu/minute	×12.96	= Foot-pounds/second
Btu/minute	×0.02356	= Horsepower
Btu/minute	×0.01757	= Kilowatts
Btu/minute	×17.57	= Watts
Btu/square foot/minute	×0.1221	= Watts/square inch
Bucket (British-dry)	$\times 1.818\times 10^{4}$	= Cubic centimeters
Bushels	×1.2445	= Cubic feet
Bushels	×2,150.4	= Cubic inches
Bushels	×0.03524	= Cubic meters
Bushels	×35.24	= Liters
Bushels	×4	= Pecks
Bushels	×64	= Pints (dry)
Bushels	×32	= Quarts (dry)

C

Calories, gram (mean)	$\times 3.9685\times 10^{-3}$	= Btu (mean)
Calories/gram	×1.8	= Btu/pound

Calories/gram-mol-°C	×1.0	= Btu/pound-mol-°F
Candle/square centimeter	×3.142	= Lamberts
Candle/square inch	×0.487	= Lamberts
Centares	×1.0	= Square meters
Centigrade	×9/5+32	= Fahrenheit
Centiliter	×0.3382	= Fluid ounce (USA)
Centiliter	×0.6103	= Cubic inch
Centiliter	×2.705	= Drams
Centimeters	$\times 3.281\times 10^{-2}$	= Feet
Centimeters	×0.3937	= Inches
Centimeters	$\times 1.094\times 10^{-2}$	= Yards
Centimeters	×393.7	= mils
Centimeters of mercury	×0.1934	= Pounds/square inch
Centimeters of mercury	×0.01316	= Atmospheres
Centimeters of mercury	×0.4461	= Feet of water
Centimeters of mercury	×136	= Kilograms/square meter
Centimeters of mercury	×27.85	= Pounds/square foot
Centimeters of mercury	×0.1934	= Pounds/square inch
Centimeters/second	×1.1969	= Feet/minute
Centimeters/second	×0.03281	= Feet/second
Centimeters/second	×0.036	= Kilometers/hour
Centimeters/second	×0.1943	= Knots
Centimeters/second	×0.6	= Meters/minute
Centimeters/second	×0.02237	= Miles/hour
Chain	×792	= Inches
Chain	×20.12	= Meters
Circumference	×6.283	= Radians
Cords	×8	= Cord feet
Cord feet	×16	= Cubic feet
Coulomb	$\times 2.998\times 10^{9}$	= Statcoulombs

Coulombs	$\times 1.036\times 10^{-5}$	= Faradays
Coulombs/square centimeter	×64.52	= Coulombs/square inch
Coulombs/square centimeter	$\times 10^{4}$	= Coulombs/square meter
Coulombs/square inch	×0.155	= Coulombs/square centimeter
Coulombs/square inch	×1.550	= Coulombs/square meter
Cubic centimeters	$\times 3.531\times 10^{-5}$	= Cubic feet
Cubic centimeters	×0.06102	= Cubic inches
Cubic centimeters	$\times 10^{-6}$	= Cubic meters
Cubic centimeters	$\times 1.308\times 10^{-6}$	= Cubic yards
Cubic centimeters	$\times 2.624\times 10^{-4}$	= Gallons (USA liquid)
Cubic centimeters	×0.001	= Liters
Cubic centimeters	$\times 2.113\times 10^{-3}$	= Pints (USA liquid)
Cubic centimeters	$\times 1.057\times 10^{-3}$	= Quarts (USA liquid)
Cubic feet	×0.8036	= Bushels (dry)
Cubic feet	×28.320	= Cubic centimeters
Cubic feet	×1.728	= Cubic inches
Cubic feet	×0.02832	= Cubic meters
Cubic feet	×0.03704	= Cubic yards
Cubic feet	×7.48052	= Gallons (USA liquid)
Cubic feet	×6.232	= Gallons (imperial)
Cubic feet	×6.428	= Gallons (USA dry)
Cubic feet	×62.425	= Pounds (water)
Cubic feet	×28.32	= Liters
Cubic feet	×59.84	= Pints (USA liquid)
Cubic feet	×29.92	= Quarts (USA liquid)
Cubic feet	×0.1781	= Barrels (oil, USA)
Cubic feet/minute	×472	= Cubic centimeters/second
Cubic feet/minute	×0.1247	= Gallons/second
Cubic feet/minute	×0.4720	= Liters/second
Cubic feet/minute	×62.43	= Pounds of water/minute
Cubic feet/minute	×1.6989	= Cubic meters/hour

Cubic feet/minute	$\times 4.719\times 10^{-4}$	= Cubic meters/second
Cubic feet/second	×0.646317	= Million gallons/day
Cubic feet/second	×448.831	= Gallons/minute
Cubic feet/second	×101.94	= Cubic meters/hour
Cubic inches	×16.39	= Cubic centimeters
Cubic inches	$\times 5.787\times 10^{-4}$	= Cubic feet
Cubic inches	$\times 1.639\times 10^{-5}$	= Cubic meters
Cubic inches	$\times 2.143\times 10^{-5}$	= cubic yards
Cubic inches	$\times 4.329\times 10^{-3}$	= Gallons (U.S)
Cubic inches	×0.01639	= Liters
Cubic inches	$\times 1.061\times 10^{5}$	= Mil-feet
Cubic inches	×0.03463	= Pints (USA liquid)
Cubic inches	×0.01732	= Quarts (USA liquid)
Cubic meters	×6.290	= Barrels (USA oil)
Cubic meters	×28.38	= Bushes (dry)
Cubic meters	$\times 10^{6}$	= Cubic centimeters
Cubic meters	×35.314	= Cubic feet
Cubic meters	×61.023	= Cubic inches
Cubic meters	×1.308	= Cubic yards
Cubic meters	×264.17	= Gallons (USA liquid)
Cubic meters	×220	= Gallons (imperial)
Cubic meters	×1.000	= Liters
Cubic meters	×2.113	= Prints (USA liquid)
Cubic meters	×1.057	= Quarts (USA liquid)
Cubic meters/hour	$\times 9.810\times 10^{-3}$	= Cubic feet/second
Cubic meters/hour	×0.5886	= Cubic feet/minute
Cubic meters/hour	×4.4033	= Gallons/minute (USA)
Cubic meters/hour	×150.95	= Barrels/day
Cubic meters/hour	×3.6651	= Imperial gallons/minute
Cubic meters/hour	×35.31	= Cubic feet/hour
Cubic meters/hour	×277.8	= Cubic centimeters/second

Cubic meters/hour	×277.8	= Cubic centimeters/second
Cubic yards	$\times 7.646\times 10^{5}$	= Cubic centimeters
Cubic yards	×27	= Cubic feet
Cubic yards	×46.656	= Cubic inches
Cubic yards	×0.7646	= Cubic meters
Cubic yards	×202	= Gallons (USA liquid)
Cubic yards	×764.6	= Liters
Cubic yards	×1,615.9	= Pints (USA liquid)
Cubic yards	×807.9	= Quarts (USA liquid)
Cubic yards/minute	×0.45	= Cubic feet/second
Cubic yards/minute	×3.367	= Gallons/second
Cubic yards/minute	×12.74	= Liters/second
Cubit (from The Bible)	×21.8	= Inch
Cup	×0.5	= Pint
Cup	×16	= Tablespoon

D

Dalton	$\times 1.650\times 10^{-24}$	= Gram
Days	×1.440	= Minutes
Days	×86.400	= Seconds
Decigrams	×0.1	= Grams
Deciliters	×0.1	= Liters
Decimeters	×0.1	= Meters
Degrees (Angle)	×60	= Minutes
Degrees (Angle)	×3600	= Seconds
Degrees (Angle)	×0.01745	= Radians
Degrees (Angle)	×0.01111	= Quadrants
Degrees/second	×0.1667	= Revolutions/minute
Degrees/second	$\times 2.778\times 10^{-3}$	= Revolutions/second

Dekagrams	×10	= Grams
Dekaliters	×10	= Liters
Dekameters	×10	= Meters
Drams (apothecaries or troy)	×0.1371429	= Ounces (avoirdupois)
Drams (apothecaries or troy)	×0.125	= Ounces (troy)
Drams	×27.34375	= Grains
Drams	×1.771845	= Grams
Drams	×0.0625	= Ounces
Dyne/centimeter	×0.01	= Erg/square millimeters
Dyne/square centimeters	$\times 9.869\times 10^{-7}$	= Atmospheres
Dyne/square centimeters	$\times 2.953\times 10^{-5}$	= Inch of mercury at 0°C
Dyne/square centimeters	$\times 4.015\times 10^{-4}$	= Inch of water at 4°C
Dynes/square centimeters	$\times 10^{-6}$	= Bars
Dynes	$\times 1.020\times 10^{-3}$	= Grams
Dynes	$\times 10^{-7}$	= Joules/centimeters
Dynes	$\times 10^{-5}$	= Joules/meters (newtons)
Dynes	$\times 1.020\times 10^{-6}$	= Kilograms
Dynes	$\times 7.233\times 10^{-5}$	= Poundals
Dynes	$\times 2.248\times 10^{-6}$	= Pounds

E

Ell	×114.30	= Centimeters
Ell	×45	= Inches
Em, pica	×0.167	= Inch
Em, pica	×0.4233	= Centimeter
Erg/second	×1	= Dyne-centimeter/second
Erg/second	$\times 1.0\times 10^{-7}$	= Watt
Erg	$\times 9.480\times 10^{-11}$	= Btu
Erg	×1	= Dyne-centimeter

Erg	×1	= Dyne-centimeter
Erg	$\times 7.367\times 10^{-8}$	= Foot-pounds
Erg	$\times 1.0\times 10^{-7}$	= Joules
Expansion coefficient, °F	×1.8	= Expansion coefficient, °C

F

Fahrenheit minus 32°F	×0.5556	= Centigrade
Famm	×5.8455	= Foot, USA
Famm	×1.7814	= Meter
Faradays	×26.8	= Ampere-hour
Fathom, British	×6.08	= Feet
Fathom, British	×1.8532	= Meter
Fathom, USA	×6	= Feet
Fathom, USA	×1.8288	= Meter
Fathom, USA	×2	= Yard
Feet	×30.48	= Centimeters
Feet	$\times 1.645\times 10^{-4}$	= Miles (nautical)
Feet, USA	×0.3048	= Meters
Feet, USA	×0.3333	= Yards
Feet, USA	$\times 0.18939\times 10^{-3}$	= Miles, USA statute
Feet, USA	×12	= Inches
Feet, USA	$\times 1.2\times 10^{4}$	= Mils
Feet, USA	×0.0606	= Rod
Feet of water	×0.0295	= Atmospheres
Feet of water	×0.8826	= Inches of mercury
Feet of water	×0.03048	= Kilograms/square centimeters
Feet of water	×304.8	= Kilograms/square meter
Feet of water	×62.43	= Pounds/square foot
Feet of water	×0.4335	= Pounds/square inch

Feet/hour	×0.01666	= Feet/minute
Feet/hour	$\times 0.2777\times 10^{-3}$	= Feet/second
Feet/hour	$\times 0.1894\times 10^{-3}$	= Miles/hour
Feet/minute	×0.5080	= Centimeter/second
Feet/minute	×0.01666	= Feet/second
Feet/minute	×0.18288	= Kilometer/hour
Feet/minute	×0.009868	= Knot
Feet/minute	×0.3048	= Meter/minute
Feet/minute	×0.00508	= Meter/second
Feet/minute	×0.01136	= Mile/hour
Feet/minute	$\times 0.1894\times 10^{-3}$	= Mile/minute
Feet/second	×30.48	= Centimeters/second
Feet/second	×1.097	= Kilometers/hour
Feet/second	×0.5921	= Knots
Feet/second	×18.29	= Meters/minute
Feet/second	×0.681818	= Miles/hour
Feet/second	×0.0113636	= Miles/minute
Feet/second	×3600	= Feet/hour
Feet/second	×60	= Feet/minute
Feet/second/second	×30.48	= Centimeters/second/second
Feet/second/second	×1.097	= Kilometers/hour/second
Feet/second/second	×0.3048	= Meters/second/second
Feet/second/second	×0.6818	= Miles/hour/second
Feet/100 feet	×1	= Percent grade
Firkin	×9	= Gallon, liquid, USA
Firkin	×34.06798	= Liter
Foot-candle	×10.764	= Lumen/square meter
Foot-candle	×1	= Lumen/square foot
Foot-candle	×10.764	= Lux
Foot-candle	×1.076	= Milliphot
Foot-candle	×0.001076	= Phot

Foot-candle	× distance in feet^2	= Candlepower
Foot-Lambert	$\times 0.3425\times 10^{-3}$	= Candle/square centimeters
Foot-Lambert	×0.3183	= Candle/square foot
Foot-Lambert	×0.00221	= Candle/square inch
Foot-Lambert	×0.001076	= Lambert
Foot-Lambert	× square foot Area	= Lumen
Foot-Lambert	×1.076	= Millilambert
Foot-Lambert	$\times 0.342\times 10^{-3}$	= Stilb
Foot-pound	$\times 1.2853\times 10^{-3}$	= Btu
Foot-pound	×1.356×107	= Ergs
Foot-pound	×0.32389	= Gram-calorie
Foot-pound	$\times 5.0505\times 10^{-7}$	= Horsepower-hours, USA
Foot-pound	$\times 5.12\times 10^{-7}$	= Horsepower-hours, metric
Foot-pound	×12	= Inch-pound
Foot-pound	×1.35582	= Joule absolute
Foot-pound	×1.3554	= Joule international
Foot-pound	$\times 3.238\times 10^{-4}$	= Kilogram-calories
Foot-pound	×0.1383	= Kilogram-meters
Foot-pound	$\times 3.766\times 10^{-7}$	= Kilowatt-hours
Foot-pound	×0.001356	= Kilowatt-second
Foot-pound	×0.01338	= Liter-atmosphere
Foot-pound	$\times 0.3766\times 10^{-3}$	= Watt-hour
Foot-pound	×1.356	= Watt-second
Foot-pound/minute	×0.077118	= Btu/hour
Foot-pound/minute	$\times 1.286\times 10^{-3}$	= Btu/minute
Foot-pound/minute	×2.259×105	= Erg/second
Foot-pound/minute	×0.01666	= Foot-pound/second
Foot-pound/minute	$\times 3.066\times 10^{-5}$	= Horsepower, metric
Foot-pound/minute	$\times 3.0303\times 10^{-5}$	= Horsepower, USA
Foot-pound/minute	$\times 2.2597\times 10^{-5}$	= Kilowatt
Foot-pound/minute	×0.022597	= Watt

Foot-pound/minute	×0.02259	= Watt
Foot-pound/second	×0.0771	= Btu/minute
Foot-pound/second	×4.6263	= Btu/hour
Foot-pound/second	$\times 1.843\times 10^{-3}$	= Horsepower, metric
Foot-pound/second	$\times 1.818\times 10^{-3}$	= Horsepower, USA
Foot-pound/second	×1.356	= Joule
Foot-pound/second	$\times 1.3558\times 10^{-3}$	= Kilowatts
Foot-pound/second	×1.3558	= Watt
Fot	×0.974	= Foot, USA
Fot	×100	= Lines
Fot	×0.2969	= Meter
Fot	×10	= Turn
Foute	×1	= Foot, USA
Furlong	×6.6	= Chain, engineer
Furlong	×10	= Chain, Gunter
Furlong	×660	= Feet
Furlong	×201.168	= Meters
Furlong	×0.125	= Mile, statue, USA
Furlong	×220	= Yards
Furlong	×40	= Rods
Fuss	×0.9842	= Foot, USA
Fuss	×0.300	= Meter

G

Gallon, British, Imperial Liquid	×0.125	= Bushel, dry, British
Gallon, British, Imperial Liquid	×4546	= Cubic centimeter
Gallon, British, Imperial Liquid	×0.16046	= Cubic foot

Gallon, British, Imperial Liquid	×0.0045	= Cubic meter
Gallon, British, Imperial Liquid	×1.032	= Gallon, dry, USA
Gallon, British, Imperial Liquid	×1.20095	= Gallon, liquid, USA
Gallon, British, Imperial Liquid	×4.54596	= Kilogram
Gallon, British, Imperial Liquid	×10	= Pound, water, 62°F
Gallon, dry, USA	×0.125	= Bushel, USA
Gallon, dry, USA	×4404.92	= Cubic centimeter
Gallon, dry, USA	×0.155555	= Cubic foot
Gallon, dry, USA	×268.803	= Cubic inch
Gallon, dry, USA	×1.16365	= Gallon, liquid, USA
Gallon, dry, USA	×4.4049	= Liter
Gallon, dry, USA	×0.05	= Peck
Gallon, dry, USA	×8	= Pint
Gallon, dry, USA	×4.6546	= Quart, liquid, USA
Gallon, liquid, USA	×0.0238	= Barrel, oil
Gallon, liquid, USA	×3785.434	= Cubic centimeter
Gallon, liquid, USA	×3.785434	= Cubic decimeter
Gallon, liquid, USA	×0.13368	= Cubic foot
Gallon, liquid, USA	×231	= Cubic inch, water, 62°F
Gallon, liquid, USA	$\times 3.7854 \times 10^{-3}$	= Cubic moter
Gallon, liquid, USA	$\times 4.951 \times 10^{-3}$	= Cubic yard
Gallon, liquid, USA	×0.859365	= Gallon, dry, USA
Gallon, liquid, USA	×0.832673	= Gallon, liquid, British
Gallon, liquid, USA	×3.7853	= Liter
Gallon, liquid, USA	×8	= Pint, liquid, USA
Gallon, liquid, USA	×4	= Quart, liquid, USA
Gallon, liquid, USA	×8.3453	= Pounds, water

Gallon/hour, USA	×0.1337	= Cubic feet/hour
Gallon/hour, USA	$\times 2.228\times 10^{-3}$	= Cubic feet/minute
Gallon/hour, USA	×0.01666	= Gallons/minute
Gallon/hour, USA	$\times 2.777\times 10^{-4}$	= Gallons/second
Gallon/minute, USA	×34.2857	= Barrels/day, oil
Gallon/minute, USA	×1.42857	= Barrels/hour, oil
Gallon/minute, USA	×0.023809	= Barrels/minute, oil
Gallon/minute, USA	×192.49999	= Cubic feet/day
Gallon/minute, USA	×8.021	= Cubic feet/hour
Gallon/minute, USA	×0.13368	= Cubic feet/minute
Gallon/minute, USA	$\times 2.228\times 10^{-3}$	= Cubic feet/second
Gallon/minute, USA	×0.2271	= Cubic meters/hour
Gallon/minute, USA	×1440	= Gallons/day
Gallon/minute, USA	×60	= Gallons/hour
Gallon/minute, USA	×0.01666	= Gallons/second
Gallon/minute, USA	×5.35565	= Tonnes, long, water, 62°F/day
Gallon/minute, USA	×5.99839	= Tonnes, short, water, 62°F/day
Gallon/minute, USA	×0.06308	= Liters/second
Gallon/second, USA	×481	= Cubic feet/hour
Gallon/second, USA	×8.02	= Cubic feet/minute
Gallon/second, USA	×0.1337	= Cubic feet/second
Gallon/second, USA	×60	= Gallons/minute, USA
Gills, British	×142.07	= Cubic centimeter
Gills, British	×0.1183	= Liters
Gills, British	×0.25	= Pints, liquid
Grade	×0.0025	= Circle
Grade	×9,000	= Degree
Grade	×54	= Minute
Grade	×0.01571	= Radian

Grain	×0.01666	= Dram, apothecary
Grain	×0.03657	= Dram, avoirdupois
Grain (troy)	×1	= Grain (avdp)
Grain (troy)	×0.0648	= Grams
Grain (troy)	$\times 2.0833\times 10^{-3}$	= Ounces (troy)
Grain (troy)	$\times 2.286\times 10^{-3}$	= Ounces (avdp)
Grain/U.S. gallon	×17.118	= Parts/million
Grain/U.S. gallon	×142.86	= Pounds/million gallon
Grain/Imperial gallon	×14.286	= Parts/million
Gram	×5	= Carat
Gram	×3.858	= Carat, metric
Gram	×100	= Centigram
Gram	×0.2572	= Dram, apothecary
Gram	×0.56438	= Dram, avoirdupois
Gram	×980.665	= Dyne
Gram	×15.4324	= Grain
Gram	$\times 9.807\times 10^{-5}$	= Joules/centimeters
Gram	$\times 9.807\times 10^{-3}$	= Joules/meter (newtons)
Gram	×0.001	= Kilograms
Gram	×1000	= Milligrams
Gram	×0.03527	= Ounces, avoirdupois
Gram	×0.03215	= Ounces, troy
Gram	×0.07093	= Poundals
Gram	$\times 2.205\times 10^{-3}$	= Pounds
Grams	×1/MW	= Gram-mols
Grams/centimeter	$\times 5.6\times 10^{-3}$	= Pounds/inch
Grams/cubic centimeter	×62.43	= Pounds/cubic foot
Grams/cubic centimeter	×0.03613	= Pounds/cubic foot
Grams/liter	×58.417	= Grains/gallon, USA
Grams/liter	×8.345	= Pounds/1,000 gallons
Grams/liter	×0.062427	= Pounds/cubic foot

Grams/liter	×1.000	= Parts/million
Grams/square centimeter	×2.0481	= Pounds/square foot
Gram-calories	$\times 3.968\times 10^{-3}$	= Btu
Gram-calories	$\times 4.1868\times 10^{7}$	= Ergs
Gram-calories	×3.088	= Foot-pounds
Gram-calories	$\times 1.55856\times 10^{-6}$	= Horsepower/hours
Gram-calories	$\times 1.163\times 10^{-6}$	= Kilowatt-hours
Gram-calories	$\times 1.163\times 10^{-3}$	= Watt-hours
Gram/calories/second	×14.286	= Btu/hour
Gram-centimeters	$\times 9.29658\times 10^{-8}$	= Btu
Gram-centimeters	×980.7	= Ergs
Gram-centimeters	$\times 9.807\times 10^{-5}$	= Joules
Gram-mol	×22.414	= Liters at 0°C and 1 atm
Gross	×12	= Dozen
Gross, great	×144	= Dozen
Gross, great	×12	= Gross

H

Hand	×10.16	= Centimeter
Hand	×4	= Inch
Hand	×48	= Foot
Hand	×1,016	= Meter
Head, feet elevation, water	×0.433	= Pounds/square inch
Hectare	×2.471	= Acre
Hectare	×100	= Are
Hectare	$\times 1.07639\times 10^{5}$	= Square feet
Hectare	×0.01	= Square kilometer
Hectare	×10,000	= Square meters

Hectare	$\times 3.861\times 10^{-3}$	= Square miles
Hectare	×11,960	= Square yard
Hectogram	×100	= Gram
Hectoliter	×3.532	= Cubic feet
Hectoliter	×0.1	= Cubic meter
Hectoliter	×0.1308	= Cubic yard
Hectoliter	×26.42	= Gallon, USA
Hectoliter	×100	= Liter
Hectometer	×328.089	= Feet
Hectometer	×100	= Meter
Hectometer	×0.06214	= Mile, statute, USA
Hectometer	×109.36	= Yard
Hectowatts	×100	= Watts
Henries	×1,000	= Millihenries
Hogsheads, British	×10.114	= Cubic feet
Hogsheads, USA	×8.42184	= Cubic feet
Hogsheads, USA	×63	= Gallons, USA
Hogsheads, USA	×238.476	= Liter
Hogsheads, USA	×504	= Pint
Hogsheads, USA	×252	= Quart
Horsepower, USA	×42.44	= Btu/minute
Horsepower, USA	×33.000	= Foot-pounds/minute
Horsepower, USA	×550	= Foot-pounds/second
Horsepower, USA	×0.7457	= Kilowatts
Horsepower, USA	×1.014	= Horsepower (metric)
Horsepower, boiler	×33.479	= Btu/hour
Horsepower, boiler	×34.5	= Pounds water/hour
Horsepower, boiler	×9.803	= Kilowatts
Horsepower, electric	×0.7072	= Btu/second
Horsepower, electric	×746	= Joule/second
Horsepower, electric	×0.746	= Kilowatts

Horsepower, electric	×746	= Watts
Horsepower (metric)	×0.98632	= Horsepower (USA)
Horsepower (metric)	×542.5	= Foot-pounds/second
Horsepower, hours, USA	×2,547	= Btu
Horsepower, hours, USA	$\times 2.6845\times 10^{13}$	= Ergs
Horsepower, hours, USA	$\times 1.98\times 10^{6}$	= Foot-pounds
Horsepower, hours, USA	×641.190	= Gram-calories
Horsepower, hours, USA	×1.01387	= Horsepower-hour, metric
Horsepower, hours, USA	$\times 2{,}376\times 10^{4}$	= Inch-pounds
Horsepower, hours, USA	$\times 26.8453\times 10^{5}$	= Joule
Horsepower, hours, USA	×0.7457	= Kilowatt-hour
Horsepower-hours, metric	×2509.83	= Btu
Horsepower-hours, metric	$\times 1.9529\times 10^{6}$	= Foot-pounds
Horsepower-hours, metric	×0.98632	= Horsepower-hour, USA
Horsepower-hours, metric	×632,467	= Gram-calories
Horsepower-hours, metric	×26.4761	= Joule
Horsepower-hours, metric	×0.73545	= Kilowatt-hour
Hours	×0.0417	= Day
Hours	×60	= Minute
Hours	×0.00137	= Month
Hours	$\times 0.1142\times 10^{-3}$	= Year
Hours	$\times 5.952\times 10^{-3}$	= Week
Hundredweight (long)	×112	= Pounds
Hundredweight (long)	×0.05	= Tonnes, long
Hundredweight (short)	×1.8	= Cubic foot
Hundredweight (short)	×45.36	= Kilograms
Hundredweight (short)	×100	= Pounds
Hundredweight (short)	×0.05	= Ton, short
Hundredweight (short)	×0.04536	= Tonnes, metric
Hundredweight (short)	×0.044643	= Tonnes, long

Inch	$\times 254\times 10^{6}$	= Angstrom
Inch	$\times 2.54$	= Centimeter
Inch	$\times 0.833\times 10^{-3}$	= Chain, engineer
Inch	$\times 1.2626\times 10^{-3}$	= Chain, Gunter
Inch	$\times 0.254$	= Decemeter
Inch	$\times 0.08333$	= Foot, USA
Inch	$\times 0.0254$	= Meter
Inch	$\times 1.578\times 10^{-5}$	= Mile, statute, USA
Inch	$\times 25.4$	= Millimeters
Inch	$\times 1{,}000$	= Mils
Inch	$\times 5.05\times 10^{-3}$	= Rods
Inch	$\times 0.02778$	= Yards
Inch, mercury	$\times 0.03342$	= Atmospheres
Inch, mercury	$\times 1.133$	= Feet of water
Inch, mercury	$\times 13.61$	= Inch height, water
Inch, mercury	$\times 70.73$	= Pound/square foot
Inch, mercury	$\times 0.49116$	= Pound/square inch
Inch, mercury	$\times 0.03453$	= Kilograms/square centimeter
Inches of mercury	$\times 345.3$	= Kilograms/square meter
Inch-pound	$\times 1.07\times 10^{-4}$	= Btu
Inch-pound	$\times 0.0833$	= Foot-pound
Inch, water, 4°C	$\times 2.458\times 10^{-3}$	= Atmospheres
Inch, water, 4°C	$\times 0.07355$	= Inches of mercury
Inch, water, 4°C	$\times 2.54\times 10^{-3}$	= Kilograms/square centimeter
Inch, water, 4°C	$\times 0.5781$	= Ounces/square inch
Inch, water, 4°C	$\times 5.204$	= Pounds/square foot
Inch, water, 4°C	$\times 0.03613$	= Pounds/square inch

J

Joule	$\times 9.48\times 10^{-4}$	= Btu
Joule	$\times 10^{7}$	= Erg
Joule	×0.7376	= Foot-pound
Joule	$\times 2.389\times 10^{-4}$	= Kilogram-calorie
Joule	$\times 1.0197\times 10^{4}$	= Gram-centimeter
Joule	$\times 2.778\times 10^{-4}$	= Watt-hours

K

Kilogram	×1,000	= Grams
Kilogram	×70.93	= Poundals
Kilogram	×2.205	= Pounds (avdp)
Kilogram	$\times 9.842\times 10^{-4}$	= Tonnes, long
Kilogram	$\times 1.102\times 10^{-3}$	= Tonnes, short
Kilogram	×0.001	= Tonnes, metric
Kilocalories	×3.968	= Btu
Kilocalories	×3,088	= Foot-lbs.
Kilocalories	$\times 1.56\times 10^{-3}$	= Hp.-hrs.
Kilocalories	$\times 1.163\times 10^{-3}$	= Kilowatt-hrs.
Kilocalories/square meter/hr/°C	×0.205	= Btu/square foot/hr/°F
Kilogram-meters	$\times 9.294\times 10^{-3}$	= Btu
Kilograms/cubic meter	×0.001	= Grams/cubic centimeter
Kilograms/cubic meter	×0.06243	= Pounds/cubic foot
Kilograms/cubic meter	$\times 3.613\times 10^{-5}$	= Pounds/cubic inch
Kilograms/cubic meter	$\times 3.405\times 10^{-10}$	= Pounds/mil-foot

Kilograms/cubic meter	$\times 8.428\times 10^{-3}$	= Ton, short/cubic yard
Kilograms/meter	×10	= Gram/centimeter
Kilograms/meter	×391.983	= Gram/inch
Kilograms/meter	×0.672	= Pounds/foot
Kilograms/meter	×0.056	= Pounds/inch
Kilograms/square centimeter	×0.9678	= Atmospheres
Kilograms/square centimeter	×32.81	= Feet of water
Kilograms/square centimeter	×28.96	= Inch of mercury
Kilograms/square centimeter	×2,048	= Pounds/square foot
Kilograms/square centimeter	×14.223	= Pounds/square inch
Kilograms/square meter	$\times 9.678\times 10^{-5}$	= Atmospheres
Kilograms/square meter	$\times 3.281\times 10^{-3}$	= Feet of water
Kilograms/square meter	$\times 2.896\times 10^{-3}$	= Inches of mercury
Kilograms/square meter	×0.03937	= Inches of water
Kilograms/square meter	×0.2048	= Pounds/square foot
Kilograms/square meter	$\times 1.422\times 10^{-3}$	= Pounds/square inch
Kilometers	×3281	= Feet
Kilometers	$\times 3.937\times 10^{4}$	= Inches
Kilometers	×1,000	= Meters
Kilometers	×0.6214	= Miles (statute)
Kilometers	×0.5295	= Miles (nautical)
Kilometers	×1.094	= Yards
Kilometers/hour	×27.78	= Centimeters/second
Kilometers/hour	×54.68	= Feet/minute
Kilometers/hour	×0.9113	= Feet/second
Kilometers/hour	×0.5396	= Knots
Kilometers/hour	×16.67	= Meters/minute
Kilometers/hour	×0.6214	= Miles/hour
Kilowatts	×56.92	= Btu/minute
Kilowatts	×1.35972	= Horsepower, metric
Kilowatts	×1,341	= Horsepower, USA

Kilowatts	×1.000	= Watts
Kilowatts	$\times 4.426\times 10^{4}$	= Foot-lbs./minute
Kilowatts	×737.6	= Foot-lbs./second
Kilowatt-hrs.	$\times 2.655\times 10^{6}$	= Foot-lbs.
Kilowatt-hours	×1000	= Watt hours
Kilowatt-hours	×3,413	= Btu
Kilowatt-hours	$\times 3.6\times 10^{13}$	= Ergs
Kilowatt-hours	×1.36	= Horsepower-hour, metric
Kilowatt-hours	×1.341	= Horsepower-hour, USA
Kilowatt-hours	$\times 3.6\times 10^{6}$	= Joules
Kilowatt-hours	×860.5	= Kilogram-calories
Kilowatt-hours	$\times 3.671\times 10^{5}$	= Kilogram-meters
Kilowatt-hours	×3.53	= Pounds of water*
Kilowatt-hours	×22.75	= Pounds of water†
Kip	×1	= Kilopound
Kip	×1.000	= Pound
Knott, USA	×51.48	= Centimeter/second
Knott, USA	×6080.2	= Feet/hour
Knott, USA	×1.8532	= Kilometer/hour
Knott, USA	×30.887	= Meter/minute
Knott, USA	×1.15155	= Mile/hour (statute)
Knott, USA	×2027	= Yards/hour
*Evaporated from and at 212°F		
†Raised from 62°to 212°F		

L

League, land	×24	= Furlong
League, land	×4.828	= Kilometer

League, land	×3	= Mile
League, marine	×5.56	= Kilometer
League, marine	×3	= Mile, nautical
League, marine	×3.45	= Mile, statute
Light year	$\times 5.9\times 10^{12}$	= Miles
Light year	$\times 9.46091\times 10^{12}$	= Kilometers
Links, engineers’	×12	= Inches
Links, surveyors’	×7.92	= Inches
Links, surveyors’	×0.66	= Feet
Links, surveyors’	×0.22	= Yard
Liters	×0.02838	= Bushels, USA, dry
Liters	×100	= Centiliters
Liters	×1.000	= Cubic centimeters
Liters	×0.035316	= Cubic feet
Liters	×61.02	= Cubic inches
Liters	$\times 6.291\times 10^{-3}$	= Barrels, oil, USA
Liters	×61.027	= Cubic inches
Liters	×0.001	= Cubic meter
Liters	$\times 1.308\times 10^{-3}$	= Cubic yard
Liters	×0.2642	= Gallon, USA, liquid
Liters	×0.2199	= Gallons (Imperial)
Liters	×1.7598	= Pint, USA, dry
Liters	×2.1134	= Pint, USA, liquid
Liters	×2.202	= Pounds of water
Liters/minute	$\times 5.886\times 10^{-4}$	= Cubic foot/second
Liters/minute	$\times 4.403\times 10^{-3}$	= Gallons/second
Liters/second	×2.1186	= Cubic feet/minute
Lumen	×0.07958	= Candlepower
Lumen	$\times 1.47\times 10^{-3}$	= Watt
Lumens/square foot	×1	= Foot-candles
Lux	×0.0929	= Foot-candles

M

Maas	×1.5	= Liter
Meter	$\times 10^{10}$	= Angstrom units
Meter	×100	= Centimeter
Meter	×3.2808	= Feet, USA
Meter	×0.01	= Hectometer
Meter	×39.37	= Inches
Meter	×0.001	= Kilometer
Meter	$\times 5.396\times 10^{-4}$	= Miles, nautical
Meter	$\times 6.214\times 10^{-4}$	= Miles, statute
Meter	×1000	= Millimeters
Meter	×1.094	= Yards
Meter	×1.179	= Vara
Meters/minute	×1.667	= Centimeters/second
Meters/minute	×3.281	= Feet/minute
Meters/minute	×0.05468	= Feet/second
Meters/minute	×0.06	= Kilometers/hour
Meters/minute	×0.03238	= Knots
Meters/minute	×0.03728	= Miles/hour
Meters/second	×196.8	= Feet/minute
Meters/second	×3.281	= Feet/second
Meters/second	×3.6	= Kilometers/hour
Meters/second	×0.06	= Kilometers/minute
Meters/second	×2.237	= Miles/hour
Meters/second	×0.03728	= Miles/minute
Microns	$\times 39.37\times 10^{-6}$	= Inches
Microns	$\times 1\times 10^{-6}$	= Meters
Micron	×0.0001	= Centimeter

Micron	×1000	= Millimicron
Mile, USA, nautical	×6,080.2	= Feet, USA
Mile, USA, nautical	×6,080	= Feet, British
Mile, USA, nautical	×72,962.5	= Inches
Mile, USA, nautical	×1.853	= Kilometer
Mile, USA, nautical	×0.333	= League
Mile, USA, nautical	×1,853.248	= Meter
Mile, USA, nautical	×1.15155	= Mile, USA, statute
Mile, USA, nautical	×2,026.73	= Yard
Mile, USA, statute	×5.280	= Feet, USA
Miles, USA, statute	×8	= Furlongs
Miles, USA, statute	×63.360	= Inches
Miles, USA, statute	×1.60935	= Kilometer
Miles, USA, statute	×8,000	= Link
Miles, USA, statute	×1,609.35	= Meters
Miles, USA, statute	×0.8684	= Mile, USA, nautical
Miles, USA, statute	×1,900.8	= Vara
Miles, USA, statute	×1.706	= Yard
Miles/hour	×44.7	= Centimeters/second
Miles/hour	×88	= Feet/minute
Miles/hour	×1.467	= Feet/second
Miles/hour	×1.609	= Kilometers/hour
Miles/hour	×0.02682	= Kilometers/minute
Miles/hour	×0.8684	= Knots
Miles/hour	×26.82	= Meters/minute
Miles/hour	×0.4470	= Meters/second
Miles/hour	×0.01667	= Miles/minute
Miles/minute	×5,280	= Feet/minute
Miles/minute	×316,800	= Feet/hour
Miles/minute	×88	= Feet/second
Miles/minute	×60	= Miles/hour

Miles/minute	×1.609	= Kilometers/minute
Miles/minute	×0.8684	= Knots/minute
Millimeter	×0.1	= Centimeter
Millimeter	$\times 3.281\times 10^{-3}$	= Feet
Millimeter	$\times 0.03937\times 10^{-2}$	= Inches
Millimeter	$\times 10^{-6}$	= Kilometers
Millimeter	×0.001	= Meters
Millimeter	$\times 6.214\times 10^{-7}$	= Miles
Millimeter	×39.37	= Mils
Millimeter	$\times 1.094\times 10^{-3}$	= Yards
Million gallons/day	×1.54723	= Cubic feet/second
Mils	$\times 2.540\times 10^{-3}$	= Centimeters
Mils	$\times 8.333\times 10^{-5}$	= Feet
Mils	×0.001	= Inches
Mils	$\times 2.540\times 10^{-8}$	= Kilometers
Mils	$\times 2.778\times 10^{-5}$	= Yards

N

Nail	×2.5	= Inch
Nepers	×8.686	= Decibels
Newton	$\times 1\times 10^{5}$	= Dynes

O

Ounces (avdp)	×16	= Drams
Ounces (avdp)	×437.5	= Grains
Ounces (avdp)	×28.349527	= Grams

Ounces (avdp)	×0.0625	= Pounds
Ounces (avdp)	×0.9115	= Ounces, troy
Ounces	$\times 2.79\times 10^{-5}$	= Tonnes, long
Ounces	$\times 2.835\times 10^{-5}$	= Tonnes, metric
Ounces, fluid	×1.805	= Cubic inches
Ounces, fluid	×0.02957	= Liters
Ounces, troy	×480	= Grains
Ounces, troy	×31.103481	= Grams
Ounces, troy	×1.09714	= Ounces, avoirdupois
Ounces, troy	×0.08333	= Pounds, troy
Ounces/square inch	×0.0625	= Pounds/square inch

P

Parsec	$\times 19\times 10^{12}$	= Miles, USA, statute
Parsec	$\times 3.084\times 10^{13}$	= Kilometers
Parts/million	×0.05833	= Grains/gallon, USA
Parts/million	×0.07016	= Grains/gallon, British
Parts/million	×8.345	= Pounds/million gallons, USA
Peck, British	×554.6	= Cubic inches
Peck, British	×2	= Gallons, British
Peck, British	×9.0919	= Liters
Peck, USA	×0.25	= Bushels
Peck, USA	×537.605	= Cubic inches
Peck, USA	×8.809582	= Liters
Peck, USA	×8	= Quarts, dry
Peck, USA	×9.3092	= Quarts, liquid
Pennyweights, troy	×24	= Grains
Pennyweights, troy	×0.05	= Ounces, troy

Pennyweights, troy	×1.55517	= Grams
Pennyweights, troy	$\times 4.1667\times 10^{-3}$	= Pounds, troy
Percent	$\times 10^{4}$	= Parts/million
Pfund, Germany	×500	= Gram
Pint, USA, dry	×0.015625	= Bushel
Pint, USA, dry	×550.6136	= Cubic centimeter
Pint, USA, dry	×0.01945	= Cubic feet
Pint, USA, dry	×33.6	= Cubic inches
Pint, USA, dry	×2	= Cup
Pint, USA, dry	×0.125	= Gallon, USA, dry
Pint, USA, dry	×0.145545	= Gallon, USA, liquid
Pint, USA, dry	×0.5506	= Liter
Pint, USA, dry	×0.0625	= Peck
Pint, USA, dry	×0.5	= Quart, USA, dry
Pint, USA, dry	×0.58182	= Quart, USA, liquid
Pint, USA, liquid	×437.2	= Cubic centimeters
Pint, USA, liquid	×0.01671	= Cubic feet
Pint, USA, liquid	×28.875	= Cubic inch
Pint, USA, liquid	×2	= Cup
Pint, USA, liquid	×0.1074	= Gallon, USA, dry
Pint, USA, liquid	×0.125	= Gallon, USA, liquid
Pint, USA, liquid	×4	= Gill
Pint, USA, liquid	×0.4732	= Liters
Pint, USA, liquid	×16	= Ounces
Pint, USA, liquid	×0.5	= Quarts, USA, liquid
Pint, USA, liquid	×0.42968	= Quarts, USA, dry
Pint, USA, liquid	×128	= Dram, fluid
Poise	×100	= Centipoise
Pole	×16.5	= Feet
Pole	×5.0292	= Meter
Pole	×1	= Rod

Pole	×5.5	= Yard
Ponce	×2.71	= Centimeter
Pood	×1.000	= Cubic inch
Pood	×40	= Funt
Pood	×4.32	= Gallon, USA
Pood	×16.3805	= Kilogram
Poundals	×13.826	= Dynes
Poundals	×14.098	= Grams
Poundals	$\times 1.383\times 10^{-3}$	= Joules/centimeter
Poundals	×0.1383	= Joules/meter
Poundals	×0.0141	= Kilograms
Poundals	×0.1383	= Newton
Poundals	×0.03108	= Pound-force
Pound-mol	×359.05	= Cubic feet at 32°F and 1 atm
Pounds	×1/Mol. Wt.	= Pound-mols
Pounds	×2267.9616	= Carats
Pounds (avdp)	×256	= Drams
Pounds (avdp)	×7,000	= Grains
Pounds (avdp)	×453.5924	= Grams (metric)
Pounds	×0.04448	= Joules/centimeters
Pounds (avdp)	×0.4536	= Kilograms
Pounds (avdp)	×16	= Ounces
Pounds (avdp)	×14.5833	= Ounces, troy
Pounds	×32.174	= Poundals
Pounds (avdp)	×1.21528	= Pounds, troy
Pounds	$\times 4.464\times 10^{-4}$	= Tons, long
Pounds	$\times 4.536\times 10^{-4}$	= Tonnes, metric
Pounds	$\times 5\times 10^{-4}$	= Tonnes, short
Pounds, troy	×5,760	= Grains
Pounds, troy	×373.24177	= Grams
Pounds, troy	×13.1657	= Ounces, avoirdupois

Pounds, troy	×12	= Ounces, troy
Pounds, troy	×240	= Pennyweights, troy
Pounds, troy	×0.822857	= Pounds, avoirdupois
Pounds, troy	$\times 3.6735\times 10^{-4}$	= Tonnes, long
Pounds, troy	$\times 3.7324\times 10^{-4}$	= Tonnes, metric
Pounds, troy	$\times 4.1143\times 10^{-4}$	= Tonnes, short
Pounds of water	×0.01602	= Cubic feet
Pounds of water	×27.68	= Cubic inches
Pounds of water	×0.1198	= Gallons
Pounds of water/minute	$\times 2.67\times 10^{-4}$	= Cubic feet/second
Pounds/cubic foot	×0.01602	= Grams/cubic centimeters
Pounds/cubic foot	×16.02	= Kilograms/cubic meter
Pounds/cubic foot	$\times 5.787\times 10^{-4}$	= Pounds/cubic inch
Pounds/cubic foot	×27	= Pounds/cubic yard
Pounds/cubic inch	×27.68	= Grams/cubic centimeter
Pounds/cubic inch	$\times 2.768\times 10^{4}$	= Kilograms/cubic meter
Pounds/cubic inch	×1,728	= Pounds/cubic foot
Pounds/cubic inch	×46,656	= Pounds/cubic yard
Pounds/hour	$\times 10.714\times 10^{-3}$	= Tonnes/day, long
Pounds/hour	$\times 12\times 10^{-3}$	= Tonnes/day, short
Pounds/hour	$\times 10.886\times 10^{-3}$	= Tonnes/day, metric
Pounds/hour	×0.45359	= Kilograms/hour
Pounds/square foot	$\times 4.725\times 10^{-4}$	= Atmospheres
Pounds/square foot	×0.01602	= Feet of water
Pounds/square foot	×0.01414	= Inches of mercury
Pounds/square foot	×4.8824	= Kilograms/square meter
Pounds/square foot	×0.1111	= Ounce/square inch
Pounds/square foot	×0.107638	= Pound/square centimeter
Pounds/square foot	$\times 6.944\times 10^{-3}$	= Pound/square inch
Pounds/square foot	×10.76387	= Pound/square meter
Pounds/square inch	×0.068046	= Atmospheres

Pounds/square inch	×2.307	= Feet of water
Pounds/square inch	×27.7	= Inch of water
Pounds/square inch	×2.036	= Inch of mercury
Pounds/square inch	×0.0703	= Kilogram/square centimeter
Pounds/square inch	×703.1	= Kilogram/square meter
Pounds/square inch	×51.714	= Millimeters of mercury
Pounds/square inch	×2.304	= Ounce/square foot
Pounds/square inch	×144	= Pound/square foot

Q

Quadrant	×0.25	= Circumference
Quadrant	×90	= Degrees
Quadrant	×5.400	= Minutes
Quadrant	×1.571	= Radians
Quarts, USA, dry	×0.03125	= Bushel
Quarts, USA, dry	×1,101.2	= Cubic centimeter
Quarts, USA, dry	×0.03889	= Cubic foot
Quarts, USA, dry	×67.20	= Cubic inches
Quarts, USA, dry	×1.1012	= Liter
Quarts, USA, dry	×1.16365	= Quart, USA, liquid
Quarts, USA, liquid	×946.331	= Cubic centimeter
Quarts, USA, liquid	×0.03342	= Cubic foot
Quarts, USA, liquid	×57.75	= Cubic inches
Quarts, USA, liquid	$\times 9.464\times 10^{-4}$	= Cubic meters
Quarts, USA, liquid	$\times 1.238\times 10^{-3}$	= Cubic yard
Quarts, USA, liquid	×4	= Cup
Quarts, USA, liquid	×256	= Dram fluid
Quarts, USA, liquid	×0.25	= Gallons
Quarts, USA, liquid	×0.946331	= Liter

Quarts, USA, liquid	×0.946331	= Liter
Quarts, USA, liquid	$\times 5.9523\times 10^{-3}$	= Oil, barrel
Quarts, USA, liquid	×32	= Ounces
Quarts, USA, liquid	×2	= Pint
Quarts, USA, liquid	×0.8594	= Quart, USA, dry

R

Radians	×57.3	= Degrees
Radians	×3,438	= Minutes
Radians	×0.6366	= Quadrants
Radians	$\times 2.063\times 10^{5}$	= Seconds
Revolutions/minute	×6.0	= Degrees/second
Revolutions/minute	×0.1047	= Radians/second
Revolutions/minute	×0.01667	= Revolutions/second
Rod	×0.165	= Chain, engineer
Rod	×0.25	= Chain, Gunters
Rod	×16.5	= Foot
Rod	×0.025	= Furlong
Rod	×198	= Inch
Rod	×25	= Link
Rod	×5.029	= Meter
Rod	×5.5	= Yard

S

Seconds, angle	$\times 2.778\times 10^{-4}$	= Degrees
Seconds, angle	$\times 16.67\times 10^{-3}$	= Minutes
Seconds, time		= Hour

Seconds, time	$\times 2.777\times 10^{-4}$	= Hour
Seconds, time	×0.166	= Minutes
Slugs	×14.59	= Kilograms
Slugs	×32.17	= Pounds
Slugs/cubic foot	×0.5154	= Gm/cubic centimeter
Snow, cubic foot	×7.2	= Pounds, 32°F
Snow, inch deep	×0.1	= Inch, water
Square centimeter	$\times 1.076\times 10^{-3}$	= Square foot
Square centimeter	×0.155	= Square inch
Square centimeter	×0.0001	= Square meter
Square centimeter	$\times 3.861\times 10^{-11}$	= Square miles
Square centimeter	×100	= Square millimeters
Square centimeter	$\times 1.196\times 10^{-4}$	= Square yards
Square feet, USA	$\times 2.296\times 10^{-5}$	= Acre
Square feet, USA	$\times 9.29\times 10^{-4}$	= Are
Square feet, USA	×929.034	= Square centimeters
Square feet, USA	×144	= Square inches
Square feet, USA	×0.0929	= Square meter
Square feet, USA	$\times 3.587\times 10^{-8}$	= Square miles
Square feet, USA	$\times 9.29\times 10^{4}$	= Square millimeters
Square feet, USA	×0.1111	= Square yards
Square inches	×6.452	= Square centimeters
Square inches	$\times 6.944\times 10^{-3}$	= Square feet
Square inches	×645.2	= Square millimeters
Square inches	$\times 7.716\times 10^{-4}$	= Square yard
Square kilometer	×247.1	= Acre
Square kilometer	×100	= Hectare
Square kilometer	$\times 10.76\times 10^{6}$	= Square feet
Square kilometer	$\times 1.55\times 10^{9}$	= Square inches
Square kilometer	$\times 10^{6}$	= Square meters
Square kilometer	×0.3861	= Square mile, USA

Square kilometer	$\times 1.196\times 10^{6}$	= Square yards
Square meters	$\times 2.471\times 10^{-4}$	= Acre
Square meters	$\times 0.01$	= Are
Square meters	$\times 0.0001$	= Hectare
Square meters	$\times 10{,}000$	= Square centimeters
Square meters	$\times 10.764$	= Square feet
Square meters	$\times 1.550$	= Square inches
Square meters	$\times 3.861\times 10^{-7}$	= Square miles
Square meters	$\times 1.196$	= Square yards
Square miles	$\times 640$	= Acre
Square miles	$\times 259$	= Hectare
Square miles	$\times 27.88\times 10^{6}$	= Square feet
Square miles	$\times 2.59$	= Square kilometers
Square miles	$\times 2.59\times 10^{6}$	= Square meters
Square miles	$\times 3.098\times 10^{6}$	= Square yards
Square millimeters	$\times 0.01$	= Square centimeters
Square millimeters	$\times 1.076\times 10^{-5}$	= Square feet
Square millimeters	$\times 1.55\times 10^{-3}$	= Square inches
Square rods	$\times 0.00625$	= Acre
Square rods	$\times 272.25$	= Square feet
Square rods	$\times 25.293$	= Square meter
Square rods	$\times 30.25$	= Square yard
Square vara	$\times 7.716$	= Square feet
Square yard	$\times 2.066\times 10^{-4}$	= Acres
Square yard	$\times 8361$	= Square centimeter
Square yard	$\times 9$	= Square feet
Square yard	$\times 1{,}296$	= Square inches
Square yard	$\times 0.8361$	= Square meters
Square yard	$\times 3.228\times 10^{-7}$	= Square miles
Square yard	$\times 8.361\times 10^{5}$	= Square millimeters

Square yard	×0.03306	= Square rods
Stone	×14	= Pound
Stone	×6.35	= Kilogram

T

Tablespoon	×0.0625	= Cup
Tablespoon	×3	= Teaspoon
Teaspoon	×0.0208	= Cup
Teaspoon	×0.333	= Tablespoon
Temperature, °C + 17.78	×1.8	= °F
Temperature, °F − 32	×0.5556	= °C
Ton, long	×1,016	= Kilogram
Ton, long	×2,240	= Pounds
Ton, long	×1.016	= Metric tonnes
Ton, long	×1.12	= Short tonnes
Ton, metric	×7.454	= Barrel, oil, 36 API
Ton, metric	×1,000	= Kilograms
Ton, metric	×2,205	= Pounds
Ton, metric	×0.9842	= Ton, long
Ton, metric	×1.1023	= Ton, short
Ton, shipping, USA	×40	= Cubic feet
Ton, shipping, USA	×2.8317	= Cubic meter
Ton, shipping, USA	×1.050	= Ton, shipping, British
Ton, short	×40	= Cubic feet
Ton, short	×268.8	= Gallons, USA, liquid
Ton, short	×4	= Hogshead
Ton, short	×907.18486	= Kilograms
Ton, short	×1,000	= Liter
Ton, short	×32,000	= Ounces

Ton, short	×2,000	= Pounds
Ton, short	×0.89286	= Tonnes, long
Ton, short	×0.907	= Tonnes, metric
Tonnes, short/square foot	×9,765	= Kilograms/square meter
Tonnes, short/square foot	×2,000	= Pounds/square inch
Tonnes, short/day	×83.333	= Pounds/hour
Tonnes, short/day	×0.16643	= Gallons/minute
Tonnes, short/day	×0.9072	= Tonnes, metric/day
Tonnes, short/day	×0.8929	= Tonnes, long/day
Tonnes, short/day	×37.8	= Kilograms/hour
Tonnes, metric/day	×91.859	= Pounds/hour
Tonnes, metric/day	×41.667	= Kilograms/hour
Tonnes, metric/day	×0.9843	= Tonnes, long/day
Tonnes, metric/day	×1.1023	= Tonnes, short/day
Tonnes, long/day	×1.12	= Tonnes, short/day
Tonnes, long/day	×1.016	= Tonnes, metric/day
Tonnes, long/day	×93.333	= Pounds/hour
Tonnes, long/day	×42.335	= Kilograms/hour

V

Vara	×2.7777	= Feet
Vara	×33.3333	= Inch
Vara	×0.9259	= Yard
Volt/inch	×0.3937	= Volt/centimeter

W

Water, 62°F, Gallon	×8.3211	= Pound

Water, 62°F, Gallon	×8.3311	= Pound
Water height in feet	×0.4335	= Pound/square inch
Water height in feet	×0.03048	= Kilograms/square centimeters
Water height in inches	×0.03613	= Pound/square inch
Water height in inches	×0.00254	= Kilograms/square centimeter
Water height in meters	×1.42067	= Pound/square inch
Water height in meters	×0.100	= Kilograms/square centimeters
Watts	×3.4128	= Btu/hour
Watts	×0.05688	= Btu/minute
Watts	×107	= Ergs/second
Watts	×44.27	= Foot-pounds/minute
Watts	×0.7378	= Foot-pounds/second
Watts	$\times 1.341\times 10^{-3}$	= Horsepower, USA
Watts	$\times 1.36\times 10^{-3}$	= Horsepower, metric
Watts	×0.001	= Kilowatt
Watts	×1	= Joules/second
Watts (abs.)	×0.056884	= Btu (mean)/minute
Watt-hours	×3.4128	= Btu
Watt-hours	$\times 3.60\times 10^{10}$	= Ergs
Watt-hours	×2,656	= Foot-pounds
Watt-hours	×858.85	= Gram-calories
Watt-hours	$\times 1.341\times 10^{-3}$	= Horsepower-hours, USA
Watt-hours	$\times 1.3596\times 10^{-3}$	= Horsepower-hours, metric
Watt-hours	×0.8605	= Kilogram-calories
Watt-hours	×367.2	= Kilogram-meters
Watt-hours	×0.001	= Kilowatt-hours

Y

Yard, USA	×91.4402	= Centimeter

Yard, USA	×3	= Feet
Yard, USA	×36	= Inch
Yard, USA	$\times 9.144\times 10^{-4}$	= Kilometer
Yard, USA	×0.9144	= Meter
Yard, USA	$\times 4.934\times 10^{-4}$	= Mile, nautical, USA
Yard, USA	$\times 5.682\times 10^{-4}$	= Mile, Statute, USA
Yard, USA	×914.402	= Millimeters
Yard, USA	×0.1818	= Rod
Year	×8,765	= Hours
Year	×525.948	= Minutes

Appendix E. Equations and Analysis

E.1. Fitting Equations to Data

Engineers, technologists, and scientists frequently collect paired data in order to understand the behavior of the system. Numerical methods of analysis have been employed to solve a wide range of steady and transient state problems. The fundamentals are essential in the basic operations of curve fitting, approximation, interpolation, numerical solutions of simultaneous linear and nonlinear equations, numerical differentiation, and integration. These requirements are greater when new processes are designed.

Various types of software packages are now readily available for scientists and engineers. They must fit a function or functions to measure data that fluctuate, which result from random error of measurement. A Fortran computer program (PROG1) was developed which determines the coefficients that provide the best fit for the following equations:

(E-1)

(E-2)

(E-3)

(E-4)

(E-5)

(E-6)

(E-7)

(E-8)

The nonlinear Eqs. (E-5), (E-6) and (E-8) can be transformed by linearizing as follows:

(E-9)

(E-10)

(E-11)

E.2. Linear Regression Analysis

Regression analysis uses statistical and mathematical methods to analyze experimental data and to fit mathematical models to these data. The least squares provide the best method for objectively determining the best straight line through a series of points. The method assumes that all deviations from the line are the result of error in the measurement of the dependent variable. The method of least squares yields the parameters which minimize the sum of squares of the residuals (e.g., the deviation of each measurement of the dependent variable from its calculated value). If F is the calculated value and *Y* is the original value of the dependent variable, then the residual sum of the squares can be minimized by the following expression.

(E-12)

where n = number of observations of X–Y data.

(E-13)

(E-14)

and

(E-15)

The values of a and b are found by minimizing Eq. (E-15). This involves taking the partial derivative of the equation with respect to each variable a and b, and setting the result to zero.

(E-16)

Substituting Eq. (E-15) into Eq. (E-16) gives

(E-17)

and

(E-18)

This is equivalent to

(E-19)

and

(E-20)

Since *b*, *X*, and Y are not functions of *a*, and the partial derivative of *a* with respect to itself is unity, Eq. (E-19) reduces to

(E-21)

Similarly, *a*, *X*, and *Y* are not functions of *b*. Therefore, Eq. (E-20) becomes

(E-22)

where a and b are constants. Equations (E-17) and (E-18) are expressed as:

(E-23)

and

(E-24)

Equations (E-23) and (E-24) apply only when fitting a straight line to a set of X-Y data points. Both equations are linear in X, Y, and n and the unknowns a and b. Using Cramer's rule for the simultaneous equations, we have

(E-25)

and

(E-26)

Solving Eqs. (E-25) and (E-26) gives

(E-27)

and

(E-28)

respectively, where all the sums are taken over all experimental observations. The developed computer program PROG1 computes the constants a and b for Eqs. (E-1) and (E-8). The program evaluates the correlation coefficient, which gives a measure of the accuracy of fit. Alternatively, a table with column heading as X_i, Y_i, X_i^2, and X_iY_i can be constructed using the Microsoft Excel spreadsheet; the constants a and b can be evaluated from Eqs. (E-27) and (E-28) respectively. The Excel spreadsheet also employs a *trendline* through the data after being plotted into a worksheet. The following are the steps being employed to fit a straight line to a set of data in the Excel spreadsheet.

1. Open a new worksheet and enter the X-data (the independent variable) in the leftmost column.
2. Enter the Y-data (the dependent variable) in the next column.
3. Plot the data as a line graph (i.e., an X–Y chart) with arithmetic coordinates. *Do not interconnect the individual data points.*
4. Activate the graph for editing. Then click on one of the plotted data points, thus selecting the data set as the active editing object (the data points will appear highlighted when this step is carried out correctly).
5. Choose *Trendline* from the Insert menu. They specify the type of curve, and request any pertinent options when the *Trendline* dialog box appears. These are, displaying the intercept, equation of the curve and its associated correlation coefficient (r^2) squared. Extrapolating the curve fit forward, that is beyond the rightmost data point or backward beyond the leftmost data point, is another option to explore. (Generally the equation of the curve and its associated r^2 value are requested for display on the graph.)
6. Press the OK button. The curve fitting will then be carried out and the results are displayed automatically.

Commercial curve fitting packages are now available in the market and details can be retrieved from the world wide web (www) using the Google search engine by typing *linear regression.* Typical websites are:

- www.ebicom.net/~dhyams
- www.analyse-it.com
- www.wessa.net.

Table E-1 shows the computer results of Example E-1.

The computer results give the constants A and B from the model Eq. (E-6) as:

where

The pre-exponential factor, S^{-1}

The Excel spreadsheet Example E-1.xls shows another example using the Lineweaver-Burk equation to determine the constants k_m and V_{max}.

E.3. Polynomial Regression

Some engineering data are often poorly represented by a linear regression. Some form of nonlinear regression can be developed, if the dependence of Y on X is known,

Figure E-1.

The Excel spreadsheet snapshot of Example E-1.

Table E-1. Curve Fitting for Two Variables

X	***−1/X***	***Y***
283.0	−0.00353	2.08
288.7	−0.00346	4.57
294.5	−0.003396	8.24
300.5	−0.003328	15.8
306.2	−0.003266	26.4
311.6	−0.00321	46.1
318.7	−0.003138	93.5

although the total convergence of this iterative regression procedure cannot be guaranteed. However, if the form of dependence is unknown, then *Y* can be treated as a general function of *X* by trigonometric terms (Fourier analysis) or polynomial function. The least squares procedure can be readily extended to fit the data to an *n*th-degree polynomial:

(E-29)

where are constants.

For this case, the sum of the squares of the residuals is minimized.

(E-30)

At the minimum, all the partial derivatives with respect to the chosen constants are zero, that is

(E-31)

This gives a system of linear equations in unknowns,

(E-32)

The above equations are set to equal zero, and can be rearranged in the following set of normal equations.

(E-33)

Equation (E-33) in matrix form becomes

Linear equations generated by polynomial regression can be ill-conditioned when the coefficients have very small and very large numbers. This results in smooth curves that fit poorly. Ill-conditioning often happens if the degree of polynomial is large and if the *Y* values cover a wide range. The error of the polynomial regression by a standard error of the estimate is shown below:

(E-34)

where

n = the degree of polynomial

N = the number of data points.

The coefficient of determination can be expressed as:

(E-35)

The correlation coefficient is given by

(E-36)

The numerator of Eq. (E-34) should continually decrease as the degree of the polynomial is raised. Alternatively, the denominator of Eq. (E-34) causes a^2 to increase once there is departure from the optimum degree.

E.4. Multiple Regression Analysis

Inadequate results are sometimes obtained with a single independent variable. This shows that one independent variable does not provide enough information to predict the corresponding value of the dependent variable. We can approach this problem, if we use additional independent variables, and develop a multiple regression analysis to achieve a meaningful relationship. Here, we can employ a linear regression model in cases where the dependent variable is affected by two or more controlled variables.

Figure E-2.

The Excel spreadsheet snapshot of Example E-2

The linear multiple regression equation is expressed as:

(E-37)

where

Y = the dependent variable

$X_1, X_2 ..., X_K$ = the independent variable
$C_0, C_1, C_2 ... ,C_K$ = the unknown regression coefficients
K = the number of independent variables.

The unknown coefficients are estimated based on n observation for the dependent variable Y, and for each of the independent variables

’s where

These observations are of the form:

(E-38)

where

Y_j = the jth observation of the dependent variable
$X_{1j},..., X_{kj}$ = the jth observation of the $X_1, X_2, ... , X_k$ independent variables.

We can use a least squares technique to calculate estimates of by minimizing the following equation:

(E-39)

Taking the partial derivatives of S with respect to , that is and setting them equal to zero, we obtain the following set of equations:

(E-40)

Equation (E-40) can be expressed in matrix form as:

where

(E-41)

(E-42)

U is a symmetric matrix.

We can obtain estimates for the coefficients by successive elimination or by solving for the inverse of U. That is,

(E-43)

where

U^{-1} = the inverse of U.

After solving for , the estimates of the dependent variable observations can be obtained as follows:

(E-44)

The power equations have often been derived to calculate the parameters of experimental data. Such an equation can be expressed in the form:

(E-45)

Table E-3. Analysis of Variance Table for Linear Multiple Regression

Source of Variance	Degree of Freedom	Sum of Squares
Total		
Regression		
Error		

We can calculate the coefficients of the independent variables, if Eq. (E-45) is linearized by taking its natural logarithm to give

(E-46)

The coefficients [illegible] can then be obtained by Gaussian elimination. Table E-3 shows the variance table for linear multiple regression. The coefficient of determination is

(E-47)

[illegible]

and the correlation coefficient is

(E-48)

[illegible]

The test statistic is the *F*-ratio, which is defined by

[illegible]

A computer program PROG3 has been developed to determine the coefficients and correlation coefficient of a multiple regression equation.

E.5. Simultaneous Equations Using the Matrix Methods

Matrix solution of sets of linear equations are solved by Excel as follows.

First, write a set of simultaneous equations in matrix format

(E-51)

[illegible]

where *A* is the coefficient matrix, *x* is the vector of unknowns, and *b* is

the result vector. To solve this matrix equation, multiply both sides from the left by the inverse of *A:*

(E-52)

By definition, a matrix times its inverse gives the identity matrix, and the identity matrix times any vector gives the vector back. The equation above reduces to the following solution:

(E-53)

The following example uses the Fortran program PROG4 and Excel spreadsheet Prog4a.xls.

Alternatively, the Excel spreadsheet Prog4a.xls is used to solve the above example problem. The procedure is as follows.

1. Start with a new worksheet, and name it Matrix Method.
2. In cell A1, enter solving sets of equations; Matrix inversion.
3. In cell B3, enter A**x** = **b**. Enter the coefficient matrix **A** and the results vector **b**.
4. In cell A5, enter input matrix **A**.
5. In cells A6:D9, enter the constants of matrix **A**:

Cell	Constant	Cell	Constant	Cell	Constant	Cell	Constant
A6	51.3	B6	43.2	C6	56.4	D6	47.4
A7	11.3	B7	11.5	C7	15.5	D7	8.5
A8	29.4	B8	31.5	C8	22.5	D8	30.4
A9	8	B9	10.3	C9	5.6	D9	13.7

6. In cell F5, enter result vector **b**
7. In cells F6:F9, enter the constants of the result vector.
8. In cell A12, enter Inverse matrix (1/**A**)
9. Select cells A13:D16, then type
= MINVERSE (A6:D9)
and press Ctrl + Shift + Enter to insert the formula into the whole section.
10. In cell F12, enter solution vector **x** = (1/*A*)**b**.
11. Select cells F13:F16, then type
+ MMULT (A13:D16, F6:F9)
and press Ctrl + Shift + Enter to insert the formula into the whole selection.
12. Turn off the gridelines, and outline the worksheet cells as shown in Matrix Method.

Figure E-3.

The Excel spreadsheet snapshot of Example E-4.

The worksheet should now look like Matrix Method, with the solution values 0.11722, 0.378939, 0.211747, and 0.305356 for α , β , γ, and δ in cells F13:F16. This is shown in Figure E-3.

E.6. Solving Simultaneous NonLinear Equations in Excel using

Solver

Solver can be used to solve a system of simultaneous equations, which can be either linear or nonlinear. Suppose the equations are represented as

(E-54)

(E-55)

(E-56)

(E-57)

Thus, we have a system of n equations in n unknowns. We can find the values of x_l, x_2, x_3, . . . , x_n that cause each of the equations to equal zero, if the function

(E-58)

is equal to zero; that is, to find the values of x_x, x_2, x_3, . . . , x_n that cause Eq. (E-58) to equal zero. Since all the terms on the right side of Eq. (E-58) are squares, they will all be greater than or equal to zero. Hence, the only way that y can be equal to zero is that each of the individual f's should be equal to zero. Therefore, the values of x_l, x_2, x_3, … , x_n that cause y to equal zero will be the solutions to the given

system of equations. The general approach used with *Solver* is to define a target function consisting of the squares of the individual equations, as indicated by Eq. (E-58) and to then determine the values of x_1, x_2, x_3, . . . , x_n that cause the target function to equal zero. To solve a system of simultaneous equations with *Solver*, proceed as follows.

1. Enter an initial guess for each independent variable in a separate cell on the worksheet.
2. Enter the equations for and y in separate cells, expressed as Excel formulas. Within these formulas, express the unknown quantities as the addresses of the cells containing the initial guess.
3. Select *Solver* from the Tools menu.
4. When the *Solver Parameters* dialog box appears, enter the following information.

   (a) The address of the cell containing the formula for y in the Set Target Cell location.

   (b) Select value of y in the Equal to line. Then enter 0 within the associated data area (i.e., determinethe values of x_x, x_2, x_3,…, x_n that will drive the target function to zero).

   (c) Enter the range of cell addresses containing the initial values of x_1, x_2, x_3, … , x_n in the area labeled *By Changing Cells.*

   (d) If you wish to restrict the range of the independent variables, click on the **Add** button under the heading *Subject to the Constraints.* Then provide the following information within the *Add Constraint* dialog box for each of the independent variables.

(i) The cell address containing the initial value of the independent variable in the Cell Reference location.
(ii) The type of constraint (i.e., ≤ or ≥) from the pull-down menu.
(iii) The limiting value in the constraint data area.
(iv) Select **OK** to return to the *Solver Parameters* dialog box or select **Add** to add another constraint.
Note that you can always change a constraint or delete a constraint after it has been added.

(e) When all the required information has been entered correctly, select **Solve.** This will initiate the actual solution procedure.

A new dialog box labeled *Solver Results* will then appear, telling you whether or not *Solver* has been able to solve the problem. If a solution has been obtained, the desired values of the independent variables will appear in the cells that originally contained the initial values. The cell containing the target function will show a value that is zero or nearly zero.

E.7. Gauss-Seidel Iterative Method

Alternatively, Fortran computer program PROG5 employs the Gauss-Seidel method for solving the four material balance equations. The four material balance Eqs. (E-65) and (E-68) are rearranged to solve for the unknown on the diagonal position of each Eq. (E-74).

(E-65)

(E-66)

(E-67)

(E-68)

(E-74)

Table E-7 shows the computer results of PROG5.

Table E-7. Solution of Simultaneous Linear Equations by Gauss-Seidel Method

NUMBER OF LINEAR EQUATIONS: 4 MAXIMUM ITERATIONS 15 CONVERGING TOLERANCE: 0.100000E-03 THE COEFFICIENT MATRIX A(1,1) ... A(N+1,N+1) IS:				
1100.0000	0.0000	0.0000	0.0000	1000.0000
1100.0000	−1400.0000	100.0000	0.0000	0.0000
0.0000	1100.0000	−1240.0000	100.0000	0.0000
0.0000	0.0000	1100.0000	−1250.0000	0.0000
THE STARTING VECTOR X(1) ....X(N) IS:				
5.000000E-01				
5.000000E-01				
5.000000E-01				
5.000000E-01				

PROCEDURE CONVERGED AFTER 5 ITERATIONS
SOLUTION VECTOR X (1) ..... X (N) IS:
9.090909E-01
6.968780E-01
6.654189E-01
5.855687E-01

Figure E-4.

Chemical reaction with recycles in four continuous stirred tanks.

Figure E-5.

The Excel spreadsheet snapshot of Example E-5.

Figure E-6.

The Excel spreadsheet snapshot of Example E-5.

Figure E-7.
The Excel spreadsheet snapshot of Example E-5.

Figure E-8.
The Excel spreadsheet snapshot of Example E-5.

Nomenclature

F = F-ratio MSE = Error mean squares MSR = Regression mean squares

MST = Total mean squares

r = Coefficient of determination

SSE = Error sum of squares

SSR = Regression sum of squares

SST = Total sum of squares

Further Reading

1. A.K. Coker, *Fortran Programs For Chemical Process Design, Analysis and Simulation*. (1995) Gulf Publishing Co, Houston, Tx .
2. B.S. Gottfield, *Spreadsheet Tools for Engineers: Excel 5.0 Version*. (1996) The McGraw–Hill Co. Inc .
3. E.J. Billo, *Excel for Chemists – A Comprehensive Guide*. (1997) Wiley-VCH, Inc .
4. B.S. Gottfried, *Spreadsheet Tools for Engineers: Excel 2000 Version*. (2000) McGraw-Hill's Best – Basic Engineering Series and Tools .

Appendix F. Solving Equations Using Excel

F.1. Microsoft Excel Solver for NonLinear Equations

The *Solver* is an optimization package that finds a maximum, minimum, or specified value of a target cell by varying the values in one or several changing cells. It uses an iterative process, beginning with trial values of the coefficients. The value of each coefficient is changed by a suitable increment, the new value of the function is then calculated, and the change in the value of the function is used to calculate improved values for each of the coefficients. The process is repeated until the desired results are obtained. The *Solver* employs the gradient methods or the simplex method to find the optimum set of coefficients.

Constraints can be applied to the *Solver* using the Add button under the heading "Subject to the Constraints". The *Solver* is an Excel Add-in, a separate software package. To save memory, it may not automatically be opened whenever Excel is started. To install the *Solver*, choose Add-Ins from the Tools menu. Then select *Solver* Add-Ins from the resulting Add-Ins dialog box. Once the *Solver* feature has been installed, it will remain installed unless it is removed by reversing the above procedure.

To use the *Solver* to perform multiple nonlinear least squares curve fitting, the procedure is as follows:

1. Start with a worksheet containing the data (independent variable X and the dependent variable Y_{obsd}) to be fitted.
2. Add a column containing Y_{calc} values, calculated by means of an appropriate formula, and involving the X-values and one or more coefficients to be varied.
3. Add a column to calculate the square of the residual (Y_{obsd} −

Y_{calc}) for each data point.

4. Calculate the sum of squares of the residuals.

5. Use the *Solver* to minimize the sum of the squares of residuals (the target cell) by changing the coefficients of the function (the changing cells).

Since the *Solver* operates by a search routine, it will find a solution most rapidly and efficiently if the initial estimates that are provided are close to the final values. Conversely, it may not be able to find a solution if the initial estimates are far from the final values. To ensure that the *Solver* has found a global minimum rather than a local minimum, a solution is obtained using different sets of initial estimates.

The least-squares regression coefficients that are returned may be slightly different depending on the starting values being provided. The Excel spreadsheet prog6a.xls (Figures F-1a-f) uses the *Solver* to determine the outlet pressure P_2 of isothermal compressible fluid flow of Example 4-9 in the Chapter 4 of the book.

F.2. Solving Equations Using Goal Seek in Excel

Microsoft Excel provides a built-in way to perform nonlinear equations using the Goal Seek command in the Tools menu. Goal-Seek varies the value of a selected cell (the changing cell) to make the value of another cell (the target cell) reach a desired value. This feature permits rapid solutions of algebraic equations using iterative (i.e. trial and error) root-finding techniques based on a series of successive refinements derived from an initial guess. For an adiabatic compressible fluid using the Excel spreadsheet (Example 4-11.xls) in Chapter 4 of the book, the implicit adiabatic equation in cell C53 is set to zero (Figures F-2a-f). The equation is expressed by

The objective is to use Goal Seek to find the value of cell B44 that makes the function (in C53) equal to zero. The accuracy of the result depends on the magnitude of the Maximum Change parameter, which can be adjusted by choosing the Calculation tab in the Options command of the Tools menu. The default value is 0.001, and adjusting the Maximum Change parameter is critical when using Goal Seek. This allows Excel to stop iterating when the change in the result is less than the Maximum Change parameter. Therefore, the Maximum Change parameter needs to be adjusted to match the value of the function. For most calculations, the Maximum Change is set to 1E-12 or 1E-15.

The steps in using Goal Seek are as follows:

1. Select Goal Seek from the Tools menu as shown in Figure F-2a.
2. Enter C53 in the *Set cell* box.
3. Put the cursor in the *To value* box and enter the desired value, zero.
4. Put the cursor in the *By Changing cell* box and enter B44 by selecting the cell or by typing.
5. Then click on *OK* box.

After a few iteration cycles the *Status* dialog box Figure (F-2e) is displayed.

If a solution is obtained, the value of the root will appear in the cell originally containing the initial guess. The value in the cell containing the formula will show a value that is close to zero (or zero). This last value will appear within the Goal Seek status dialog box.

Cell B44 gives the root of the function cell (B44), which causes the target value to equal zero. The final value of the function depends to some extent on the starting value of Ma_1. The likelihood of obtaining a converged solution will be enhanced, if the initial guess is as close as possible to the desired root.

For problems requiring the variation of two or more parameters, that is, varying the values of several cells to make the value of another cell reach a desired value involves the use of the *Solver*.

The Excel spreadsheet Prog6a.xls uses the Goal Seek routine to determine the friction factor of Colebrook–White implicit equation as defined by

(F-1)

Figure F-1.

Nonlinear equation estimation using *Solver* from the tools menu of Microsoft Excel of Example 4.9.

Figure F-2.

Solving implicit adiabatic compressible fluid flow equation of Example 4-11 using Goal Seek.

where

ε = Relative pipe roughness, in.

D = Pipe size, in.

f_D = Darcy friction factor

Re = number.

Rearranging the friction factor equation in the form

(F-2)

The worksheet (Solution-1) in Excel spreadsheet program (Prog6a.xls) calculates the friction factor for given relative pipe roughness, pipe size, and Reynolds number. The Excel spreadsheet program Prog6a.xls and the Fortran program PROG6 show how the friction factor is determined. Table F-1 shows the computer results of a 2 in. (internal diameter = 2.067 in.) stainless steel pipe size Sch 40, with pipe roughness of 0.0018 in. and Reynolds number of 184,000 (highly turbulent flow). The calculated friction factor f = 0.02063 with

an initial guess of 0.015.

Table F-1 Newton–Raphson Method for a NonLinear Equation

Initial Guess of the root x = 0.0150		
Iteration	x	F(x)
1	0.19423E−01	0.22078E+00
2	0.20577E−01	0.93463E−02
3	0.20630E−01	0.17983E−04
Tolerance met in three iterations		
Final root x = 0.20630E−01		
F(x) = 0.17983E−04		

The worksheet (Solution-2) in Excel spreadsheet Prog6a.xls calculates the friction factor for flow of a suspension of fibrous particles to the Reynolds number given by an empirical equation of the form [5]:

(F-3)

where

f = friction factor

Re = Reynolds number

k = constant determined by the concentration of the suspension.

Rearranging Eq. (F-3) gives

(F-4)

Further Reading

1. Coker, A.K., *Fortran Programs For Chemical Process Design, Analysis and Simulation*, Gulf Publishing Co., Houston, TX, 1995.
2. Gottfield, B.S., *Spreadsheet Tools for Engineers Excel 5.0 Version*, The McGraw-Hill Co. Inc., 1996.
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4. Gottfried, B.S., *Spreadsheet Tools for Engineers Excel 2000 Version*, McGraw-Hill's Best – Basic Engineering Series and Tools, 2000.
5. Lee and Duffy, "Relationships Between Velocity Profiles and Drag Reduction in Turbulent Fiber Suspension Flow", *AIChE J.*, 1976, pp. 750–753.

Appendix G. Analytical Techniques

G.1. Useful Integrals

(G-1)

(G-2)

(G-3)

(G-4)

(G-5)

(G-6)

(G-7)

(G-8)

(G-9)

(G-10)

(G-11)

(G-12)

(G-13)

(G-14)

(G-15)

(G-16)

(G-17)

(G-18)

(G-19)

(G-20)

(G-21)

(G-22)

(G-23)

(G-24)

(G-25)

(G-26)

(G-27)

Useful Websites

- www.integrals.com
- http://integrals.wolfram.com

G.2. Liebnitz's Rule – Higher Derivatives of Products

(G-28)

where

Example

(G-29)

G.3. Definition of a Derivative

(G-30)

G.4. Product Rule

(G-31)

G.5. Quotient Rule

(G-32)

(G-33)

G.6. Exponential/Logarithmic Functions

(G-34)

(G-35)

(G-36)

G.7. Taylor's and Maclaurin's Series

Taylor's Series

(G-37)

(G-38)

when $h = 0$

Substituting into the expansion for Taylor's expression

(G-39)

and rearranging gives the following.

(G-40)

Maclaurin's Series

Maclaurin's series is a special case of Taylor's series. If $x = 0$ and h is

replaced by x, then Taylor's expansion becomes

(G-41)

G.8. Differential Equations

Separation of Variables

(G-42)

Solve by direct integration

(G-43)

Example Continuous Flow Stirred Tank Reactor

Determine C as a function of time.

Step 1 A mass balance on tank

Mass in − Mass out = Accumulation in tank

(G-44)

Step 2 Separate C and t and integrate

(G-45)

(G-46)

or

(G-47)

(G-48)

Step 3 Determine A using the boundary condition $t = 0$, $C = C_i$, therefore $A = C_i - C_O$

Step 4. The final solution is

(G-49)

or

(G-50)

Variable Separable

(G-51)

or

(G-52)

in which $F(x)$ is a function of x only, and $f(y)$ a function of y only.

The general solution is

(G-53)

Example

Solve the differential

(G-54)

dividing by xy,

(G-55)

(G-56)

(G-57)

where the constant $C_1 = \ln C$

(G-58)

giving

(G-59)

The factor used to multiply throughout to separate the variables is called an integrating factor.

G.9. Linear Equations

An equation of the form

(G-60)

is called a linear differential equation where P and Q are constants or functions of x only.

Multiplying such an equation throughout by the integrating factor gives an equation that can be solved.

(G-61)

Differentiating gives

(G-62)

which is the left-hand side of the equation.

Therefore the solution is

(G-63)

Example

Solve the equation

(G-64)

(G-65)

Multiplying the equation by this integrating factor e^{2x} gives

(G-66)

Integrating this equation gives

(G-67)

where A is a constant.

Therefore, the solution is

(G-68)

G.10. Exact Differential Equation

If M and N are functions of and then is an exact differential equation, the solution of which is

(G-69)

or

(G-70)

If the condition is not satisfied, there exists a function such that

G.11. Homogeneous Second Order Linear Differential Equation with Constant Coefficients

(G-71)

First construct and solve the auxiliary equation

(G-72)

If the auxiliary equation has distinct real roots m_1 and m_2, then the solution is

(G-73)

If the auxiliary equation has complex roots, m_1+im_2 and m_1-im_2, then the solution is:

(G-74)

If the auxiliary equation has two roots equal to m, then the solution is

(G-75)

G.12. Table of Laplace Transform

where are the values of the first, second, up to the (n−1)th derivative of the function when the independent variable is zero.

G.13. Cubic Equations

(G-76)

This reduces by the substitution

(G-77)

where

(G-78)

(G-78)

(G-79)

(G-80)

(G-81)

(G-82)

If a, b, c, and d are real and if

- $D > 0$ there are one real and two conjugate imaginary roots
- $D = 0$ there are three real roots of which at least two are equal
- $D < 0$, trigonometric formulas of the roots are

(G-83)

(G-84)

(G-85)

The value of ϕ is calculated from the expression

(G-86)

Appendix H. Numerical Techniques

This appendix shows the various numerical techniques that can be employed in solving design problems, which could pose difficulty if an analytical method is used. The numerical methods can readily be incorporated into computer programs to obtain results of design problems.

H.1. Simpson's Rule for Area Under the Curve

Simpson's rule is a numerical integration technique that is widely used in calculating the area under the curve. It is simple and has a greater degree of accuracy than the trapezoidal rule. The Simpson's 1/3 rule is based on quadratic polynomial interpolation.

Figure H-1 shows a section of a curve and three coordinates erected to it at equally spaced intervals along the x-axis. Simpson's rule states that the area P′PQQ′ is given approximately by the formula

(H-1)

If we reduce the step size *h*, the result becomes more accurate. The interval over which the integral is to be taken is divided into larger number of equal sub intervals as shown in Figure H-2.

We will divide the total area into four sections, namely P′PRR′, R′RSS′, S′STT′, and T′TQQ′.

We shall write down the expression for each area and sum them up to obtain the total area P′PQQ′.

(H-2)

(H-3)

Figure H-1.

Figure H-2.

(H-4)

(H-5)

The total area is the sum of the areas P′PRR′, R′RSS′, S′STT′, and T′TQQ′.

(H-6)

The Simpson's 1/3 rule for a quadratic integrated over two Δx intervals that are of uniform width or panel is

(H-7)

Simpson's 3/8 Rule

Simpson's 3/8 rule is derived by integrating a third-order polynomial interpolation formula. For a domain (a, *b)* divided into three intervals, it is expressed as

(H-8)

A computer program PROG7 has been developed to determine the area under the curve. Simpson's rule is also easy to use in Microsoft Excel spreadsheet. We can carry this out by entering the *x*-values in one column, the *y*-values in the next column. We then follow this procedure by an additional column containing the *y*-values multiplied by their appropriate constants (i.e., multiplied by 4 or 2, except the initial and final values). We sum the last column to obtain the value of the integral and multiply this by $\Delta x/3$.

H.2. NonLinear Equations

The Newton-Raphson's iterative method is a process for the determination of a real root of an equation $f(x) = 0$, given just one

point close to the desired root (Figure H-4). If we let x_0 represent the known approximate value of the root of $f(x) = 0$, and h be the difference between the true value α and the approximate value, we have

From Taylor series,

(H-13)

about x_0, gives

(H-14)

Figure H-4.
Newton–Raphson's method.

where

, lies between α and x_0. Ignoring the remainder term and writing $f(\alpha) = 0$, we have

H-15

so that

H-16

Therefore, the next root that gives a better estimate than x_0 is

H-17

Better approximations may be obtained by repetition (iteration) of the process. We may write this as

(H-18)

Each iteration provides the point at which the tangent at the original point cuts the x-axis as shown in Figure H-4. The equation of the tangent at the point is

(H-19)

Therefore the point (x_{n+1}, 0) corresponds to

(H-20)

which gives

(H-21)

A developed computer program PROG6.FOR is used to solve nonlinear (e.g., Colebroke-White) equation using Newton–Raphson's method of iteration in Chapter 9.

Solution of NonLinear Equations

Consider a set of N nonlinear equations of the form

(H-22)

where x, y,… are the roots of the N equations. These equations can be solved explicitly for the roots. If we consider some points (x_1, y_1) near the root of definition for the functions F, G, we can expand both functions by an N-dimensional Taylor series about the point (x_1, y_1) as

(H-23)

Truncating the series after the first-order derivative and rewriting in matrix form will yield

(H-24)

We can solve for the roots x, y, . . . which give as

(H-25)

for $N = 1$

Equation (H-25) becomes

(H-26)

This is Newton-Raphson's method for finding the roots of an equation.

When $N = 2$

(H-27)

Equation (H-27) is a two-dimensional generalization of Newton's method. This technique is often employed to solve large sets of nonlinear algebraic equations. Care must be taken in choosing initial guess (x_1, y_1) quite close to the final roots, as the algorithm may diverge. A developed computer program PROG8 is used to solve nonlinear equations as illustrated by the following example.

The computer program PROG8 can be used to solve any number of nonlinear equations. The partial derivatives of the functions are estimated by the difference quotients when a variable is perturbed by an amount equal to a small value (δ) used in the program to perturb the X-values. Table H-2 shows the computer results of Example H-2.

H.3. Solution of Simultaneous, First-Order, Ordinary Differential Equations

Analytical solutions to complex kinetic reactions in reactor systems are time-consuming and intractable. The designer must report to numerical techniques with the aid of a computer for his or her solutions. The reactions taking place in batch and piston flow reactions involve a set of simultaneous, first-order, ordinary differential equations. Several numerical methods have been used for solving sets of equations, and the most popular method is the Runge–Kutta fourth order. The Runge–Kutta is a powerful integration technique that can be easily implemented on a personal computer. The only drawback is its instability if the step size is too large. In a set of equations, the Runge–Kutta algorithm uses the same step size for each member of the set of equations. This causes practical problems if the set is *stiff* where some members of these equations have characteristic times much smaller than other members of the equations. An example is the free-radical kinetics reaction, which has rates that may differ by three orders of magnitude.

In general, a system of nth first-order equations will be of the form

(H-42)

with n initial conditions $y_i(X_0) = A_i$, $i = 0, 1, 2, \ldots, n$ Consider the system of two equations:

(H-43)

(H-44)

We may advance the solution of y and z to new values at $x_1 = x_0 + h$ using any of the one-step or Runge–Kutta methods.

Table H-2.
Newton Method
For Solving
Nonlinear
Equations

Initial roots
X(1) = 0.1000
X(2) = 0.1000
X(3) = 0.1000
X(4) = 0.1000
X(5) = 0.1000
X(6) = 0.1000
X(7) = 0.1000
X(8) = 0.1000
Final roots
X(1) = 0.54599
X(2) = 0.11372
X(3) = 0.37030
X(4) = 0.05386

X(4) = 0.05380
X(5) = 0.27180
X(6) = 0.02958
X(7) = 0.21107
X(8) = 0.01792

In general, our solutions will be advanced using expressions of the form

(H-45)

(H-46)

where the nature of *K* or *L* depends on the method being applied. For the Runge-Kutta fourth order

(H-47)

and

(H-48)

where

(H-49)

(H-50)

(H-51)

(H-52)

(H-53)

(H-54)

(H-55)

(H-56)

H.4. Extension of Runge–Kutta Methods

Runge–Kutta–Gill Method

The Runge–Kutta–Gill method is the most widely used single-step method for solving ordinary differential equations. For the differential equation

(H-57)

(H-58)

where

(H-59)

The Runge-Kutta-Merson Method

The Runge-Kutta-Merson method outlines a process for deciding the step size for better predetermined accuracy. For this method, five functions are evaluated at every step. The algorithm is

(H-60)

We can estimate the local error from a weighted sum of the individual estimate.

(H-61)

The following Example H-3 illustrates the use of Runge–Kutta fourth order with the Excel spreadsheet.

H.5. Partial Differential Equation

If two or more independent variables are involved in a differential

equation, we can express the differential equation as partial differential equation (PDE).

We shall consider the second-order equation (PDE) of the form

(H-65)

where the coefficients A, B, C, and D are functions of x, y, and u; x and y are the independent variables and u is the dependent variable We can classify Eq. (H-65) with respect to the sign of the discriminant . The equation is elliptical if $\Delta < 0$, hyperbolic if $\Delta > 0$, and parabolic type if $\Delta = 0$.

Elliptical Equation

(H-66)

and the Laplace equation

(H-67)

The coefficients $A = C = 1$, $B = 0$ and are examples of elliptical equation.

Hyperbolic Equation

The wave equation

(H-68)

is a hyperbolic type where the coefficients $A = 1$, $B = 0$ $C = -1$ and

Parabolic Equation

The heat conduction equation

(H-69)

is a parabolic type where the coefficients $A = 0$, $B = 0$, $C = 1$ and .

A method for solving the above partial differential equations is to replace the derivatives by difference quotients; that is, converting the equation to a difference equation. We can then write the difference equation that corresponds to each point at the intersections (nodes) of a grid work that subdivides the region of interest at which the function values are known. Solving these equations

Figure H-7.

simultaneously gives values for the function at each node that approximates the true values.

Let $h = \Delta x$ equal spacing of grid work in the x – direction as shown in the Figure H-7.

From Taylor series,

(H-70)

and

(H-71)

It follows that

(H-72)

or

(H-73)

Using the subscript notation, we have

(H-74)

where the subscripts of f indicate the x-values at which it is evaluated. The order relation $O(h^2)$ shows that error approaches proportionality

to h^2 as $h \to 0$.

Similarly, the first derivative is approximated to

(H-75)

or

(H-76)

(H-77)

The first derivative could also be approximated by the forward or backward difference, but would have an error of O(h). The central difference approximation gives the more accurate approximation.

When f is a function of both x and y, we can obtain the second partial derivative with respect to , by holding y constant and evaluating the function at three points where x equals , and . Correspondingly, the partial derivative is determined by holding x constant.

Consider the Laplace equation on a region in the xy plane. We subdivide the region with equispaced lines parallel to the x- and y-

axes. Consider the region near $(x_i\ y_i)$.

(H-78)

We can replace the derivatives by difference quotients which approximate the derivatives at point (x_i, y_i). Then we have

(H-79)

or

(H-80)

If we let $\Delta x = \Delta y = h$, the PDE becomes

(H-81)

This is known as the standard five-point formula, as five points are involved in the relationship of Eq. (H-81) which points to the right, left, above, and below the central point (x_t, y_t). We can write Eq. (H-81) as

(H-82)

Instead of Eq. (H-82), we may also use the formula

(H-83)

Therefore, the general procedure is to approximate the PDE by a finite difference transformation and then to obtain the solution at the mesh points, using the finite difference approximations. Other numerical methods of solution are the implicit Crank-Nicolson method or the alternating direction implicit scheme (ADI) by Peaceman and Rachford. Details of these methods are illustrated in Numerical analysis texts.

H.6. The Explicit Method

The explicit method divides space and time into discrete uniform subintervals and replaces both time and space derivatives by finite difference approximations, permitting one to easily compute values of the function at a time Δt after the initial time. These values are then used to compute a second set of values and the process is repeated. An approach to solving parabolic partial differential equations by a numerical method is to replace the partial derivatives by finite-difference approximations. An example is the one-dimensional heat flow equation.

(H-84)

We can use the relations

(H-85)

and

(H-86)

The subscripts are used to denote the position and superscripts for time.

Substituting Eqs (H-85) and (H-86) into Eq. (H-84) and solving for gives the equation for the forward-difference method:

(H-87)

Solving for in terms of the temperatures at time t_j in Eq. (H-87) for a parabolic partial-differential equation involves subdividing the length into uniform subintervals and applying the finite-difference approximation to Eq. (H-84) at each point where u is not known. Eq. (H-87) then gives the values of u at each interior point at $t = t_1$ since the values at $t = t_0$ are given by the initial conditions. It can be used to get values at t_2 using the values at t_1 as initial conditions, so the solution can be stepped forward in time. At the end points, the boundary conditions will determine u.

The relative size of the time and distance steps, Δt and Δx, affects Eq. (H-87). If the ratio of is chosen so that , the equation is simplified in that the last term vanishes and we have

(H-88)

If the value is chosen as less than one-half, there will be improved accuracy (limited by the errors dependent on the size of Δx). If the value is chosen greater than one-half, which would reduce the number of calculations required to advance the solution through a given interval of time, the phenomenon of instability sets in. An example is used to illustrate the explicit method.

H.7. Initial Value Methods

Another method for solving boundary value problems is to treat them like initial value problems. Since a second-order equation can be reduced to two first-order equations, two initial conditions are required. One condition will be known at a boundary: simply assume a value for the other dependent variable at that same boundary, integrate to the other side, and check if the required boundary condition is satisfied. If not, change the initial value and repeat the integration. The result of this method depends upon the skill with which the iterations are employed in the program.

Finite Difference Method in Excel

The Microsoft Excel spreadsheet can be used to program the finite difference method and use the “Calculation” feature to handle the circular reference. First, turn off the iteration, prepare the spreadsheet, and then turn the calculation back on. Whether this converges depends upon the initial guess.

Consider a reaction and diffusion in a flat layer with a first-order reaction represented by

(H-89)

where

The idea in the finite difference method is that the differential equation, valid for all x positions, is replaced by a set of equations representing the equation only at the grid points. Using the finite difference method as derived above, the equation at the grid point is

(H-90)

Rearranging Eq. (H-90) to the following form gives

(H-91)

Identify the cells (B1, C1, and D1) in the spreadsheet with the value of concentration at a node i as shown below.

A1	B1	C1	D
	c_{i-1}	c_i	c_{i+1}

Put the value of in cell A1. The equation for cell C1 is represented by Eq. (H-92) as

(H-92)

Next, copy this equation over a series of cells that correspond to the

number of grid points. For the first and last cell (grid point) use a different equation appropriate to the boundary condition. Then turn on the iteration feature to get the solution.

The next step is to resolve the problem using more grid point and a smaller Δx. This entails that the spreadsheet must be programmed, which gives an indication of whether the answer changes much as the mesh is refined. This ensures that the problem is accurately solved.

H.8. Finite Difference Method for Elliptic Equations

Consider the heat condition equation with a heat generation term as:

(H-93)

The finite difference form of Eq. (H-93) is solved on a worksheet by replacing the derivatives with central differences centered on grid point *(i, j)*.

(H-94)

where $T_{i,j}$ is the temperature at the ith location in the *x*-direction and the *j*th location in the y-direction. Assuming that the grid spacing , Eq. (H-94) is rearranged as

(H-95)

The spreadsheet is then arranged as shown below:

		C4 – $T_{i,j-1}$	
	B5 − $T_{i-1,j}$	C5 – $T_{i,j}$	D5 – $T_{i+1,j}$
		C6 – $T_{i,j+1}$	

When the value of is placed in cell A4, the equation for cell C5 is:

(H-96)

In the worksheet, the boundaries of the problem are set with the fixed or derivative values of the boundary conditions, and the interior points are set with Eq. (H-96). Then copy this for every internal grid point, set the boundary equations, and turn on the iteration feature to obtain the solution. This is further carried out again with a finer mesh to assess the accuracy. If the heat of generation term depends upon temperature, it is easy to include that complication just by inserting the formula in place of A4.

Further Reading

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Appendix I. Screenshot Guide to Absoft Compiler Graphical User Interface

The graphical interface to the Absoft compiler can be started from Windows by selecting Programs from the Start menu and then compile from the Absoft Tools submenu. A command line version of an individual compiler can be started with one of the commands f90, f77, or cc. Both methods allow the user to compile source files quickly and easily. The Absoft compilers can be run either from a command line or from the Absoft Developer Tools Interface.

Note: It is assumed that the compiler has been installed on the C: drive. If this is not the case, substitute the correct drive letter in the examples as appropriate.

Selecting **Programs** from the **Start** menu and then **Compiler Interface** from the Absoft submenu opens the Absoft Developer Tools Interface. It can also be started by selecting **Run** from the **Start** menu and typing c:\Absoft\bin\Atools or by typing Atools at a console or command line window. The Windows NT **Command Prompt** or Windows XP **MS-DOS Prompt** should be opened by selecting **Programs** from the **Start** menu and then **Development Command Shell** from the **Absoft** submenu. Opening the console window in this way automatically sets the environment variables by executing the c:\Absoft\bin\absvar.bat batch file.

Error may occur when compiling a program in the source file syntax or with the compiler options selected. All errors and warning messages are shown in the compiler output window in a format similar to the following:

error on line 38 of PROG1.FOR: spelling error?

The following steps as illustrated in Figure I-1Figure I-2Figure I-3Figure I-4Figure I-5Figure I-6Figure I-7Figure I-8Figure I-9Figure I-10Figure I-11 and Figure I-12 are used to compile and execute the Fortran programs in this book using the Absoft compiler.

Figure I-1.
Absoft compiler window.

Figure I-2.
New Project window from File submenu.

Figure I-3.
Absoft Fortran 77 compiler dialog box.

Figure I-4.
Source file from Configure submenu.

Figure I-5.

Source file (e.g. PROG1) is selected and added to the File(s) Added box.

Figure I-6.

f77 General Options.

Figure I-7.

f77 Compatibility Options.

Figure I-8.

Source code Window from Absoft Editor.

Figure I-9.

Compiled source code from Tools submenu.

Figure I-10.

Errors and warning messages shown in the compiler output window.

Figure I-11.
Execute the compiled source file from Tools submenu.

Figure I-12.
Results of the compiled program.

What to do	How to perform task
1. Start up the interface to the compiler: Invoke the Absoft Developer Tools interface.	Select Programs from the Start menu, then select Compiler Interface from the Absoft Pro Fortran submenu. A windows appears entitled New Project windows, select a program file name (e.g. PROG1).
2. Set the options to create an MRWE application (a Windows program).	In the target type box, click on the MRWE application button; in the Options box, click the MRWE Events box. Then click on OK box.
3. Select the source file (e.g. PROG1.For) to compile	Click on Configure menu and select Add/Remove File(s) from the pull-down menu. Select the application source file from its folder (e.g. c:\absoft60), then add it to the Source File(s) Added box by clicking on the Add button. Click on the close box.
4. Select F77 compiler options	Click on Configure menu and select Set Options from the pull-down menu.

The options fall into five categories: General, Control, Compatibility, Miscellaneous, and Format.	Using General options subset, click on check Array Boundaries. Using Compatibility subset, click on Static storage and Fold to Upper Case boxes. The Click on OK.
5. Invoke the Absoft compilers with all selected options and source files	Click on the Tools menu and Select Build from the pull-down menu. Absoft compiler successfully compiles the source file and gives a message: “Build completed”.
6. Execute the compiled application.	Click on the Tools menu and select Execute from the pull-down menu.
7. View/Print the result from Notepad, if the result file is created in the application	Click on Start menu, select programs, Accessories, Notepad in sequence. Click on file, select open from the pull-down menu. Select the absoft60 folder from the C directory. Select All files (*.*) from the File of type dialogue box. Double click on the result file (e.g. RES1). Click on Print menu to obtain a hard copy of the result.
The source codes and executable files of the various programs in the text are on the CD-ROM.	
The Absoft Compiler can be purchased from	
Absoft Corporation (www.absoft.com)	

Appendix J. Equilibrium K-Values

Figure J-1.
Pressure.vs. K for nitrogen at convergence pressure of 2000 psia (13,800 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-2.
Pressure .vs. K for ethane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-3.
Pressure .vs. K for propane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-4.

Pressure .vs. K for i-butane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-5.

Pressure .vs. K for n-butane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-6.

Pressure .vs. K for i-pentane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-7.

Pressure .vs. K for hexane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-8.

Pressure .vs. K for heptane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-9.

Pressure .vs. K for Octane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-10.

Pressure .vs. K for Nonane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Figure J-11.

Pressure .vs. K for Decane at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed.,

V. 1 and 2, (2004), Tulsa, Okla.

Figure J-12.

Pressure .vs. K for hydrogen sulfide at convergence pressure of 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2, (2004), Tulsa, Okla.

Appendix K. Simulations

Step 1. Double click on Unisim icon to display the simulation window as shown in Figure K-1.

Figure K-1.
Snapshot of Unisim design window.

Step 2. From the Tools menu, select Preferences to set up the default units as shown in Figure K-2.

Figure K-2.
Snapshot of Unisim design window.

Step 3. Select the Variables tab on the Session Preferences window.

Figure K-3.
Snapshot window of the Session Preference window.

Step 4. Set up the units as default in the Session Preferences Window. Select New User from the list in **Available Unit Set**, Select the units S.I/Field from Display Units and click on **"Save Preference As"** as

shown in Figure K-4.

Figure K-4.
Snapshot window of variables in the Session Preferences.

Step 5. Click on File menu, select New and click on Case in the menu. This allows for a new simulation problem using UniSim Design R370 software as shown in Figure K-5.

Figure K-5.
Snapshot of creating a new simulation design.

Step 6. Select the Component List View window by clicking on the tab button. Click on the Add Pure button to select the components as highlighted in Figure K-6.

Figure K-6.
Snapshot of the Component List View window showing the components.

Step 7. The Simulation Basis Manager window appears as shown in Figure K-7. Click on Fluid Pkgs tab as shown in Figure K-6. Then click on the Add button to select the components.

Figure K-7.

Snapshot window of Simulation Basis Manager.

Step 8. Click on the Add Pure button to select the components as highlighted in Figure K-8. Close the window

Figure K-8.

Snapshot of the component List View window showing the components.

Step 9. The Simulation Basis Manager window appears as shown in Figure K-9. Click on Fluid Pkgs tab as shown in Figure K-9. Then click on the **Add** button.

Figure K-9.

Snapshot of Simulation Basis Manager window.

Step 10. Choose the required property package from the **Property Package Selection** menu. For this example, choose **Peng Robinson** as shown in Figure K-10. Select **EOSs** from **Property Package Filter** and close the window.

Figure K-10.

Snapshot of Simulation Basis Manager window listing Property Package Selection.

Step 11. Close the window and return to the Simulation Basis Manager window as shown in Figure K-11. Click on **Enter Simulation Environment** button as shown in Figure K-11.

Figure K-11.
Snapshot of Simulation Basis Manager window.

Step 12. Here, the Process Flow Diagram (PFD) window appears, then from the palette, double click on the **blue** arrow and place it on the PFD window. Next, select the shortcut distillation icon in the palette by double clicking on it and place this on the PFD window as shown Figure K-12. This is named **T-100**. Double clicking on the arrow-1 gives the worksheet window.

Figure K-12.
Snapshot of the feed stream I and distillation column T-100.

Step 13. Enter the **Stream Name, Pressure, Molar flow** from **Conditions** menu in the Worksheet of the Feed window as shown in Figure K-13.

Figure K-13.
Snapshot of the Feed window showing the **Worksheet** parameters.

Step 14. Next, select Composition menu from the Worksheet in the Feed window. This provides the five components that were previously selected for this example.

Figure K-14.
Snapshot of the Feed window with the selected five components.

Step 15. Select **Mole fraction** tab from **Input Composition for Stream: Feed** window and input the data values of the components as shown in Figure K-15. Click on **OK** button to close the window.

Figure K-15.
Snapshot of **Input Composition For Stream: Feed** window.

Step 16. The total Mole fraction is 1.0 as shown in Figure K-16. Close this window and double click on the column icon on the PFD window.

Figure K-16.
Snapshot of the Feed window showing the total mole fraction of the components.

Step 17. Double clicking on the Red icon of the column shows the

Column window as shown in Figure K-17. Click on the Design tab and enter the different streams: Feed, Distillate, Bottoms, Reb-Q, Cond-Q. Choose Liquid button for Top Product Phase from Connections as shown in Figure K-17.

Figure K-17.
Snapshot of **Column** window from **Design** tab in **Connections** menu.

Step 18. Next, select the Parameter menu from Design window. Select light key in Bottoms and its Mole fraction tab, Heavy key in Distillate and its corresponding Mole fraction tab as shown in Figure K-18. Notice the highlighted red comments displayed **"Unknown Key Components"** at the bottom of the window.

Figure K-18.
Snapshot of **Column** window showing **Parameters** menu.

Step 19. Input the Parameter values and notice the highlighted red comments displayed **"Unknown Pressure"** as shown in Figure K-19.

Figure K-19.
Further snapshot of **Column** window showing **Parameters** menu.

Step 20. Next, select **Parameters** from the **Design** window. Select

Light Key in Component name from the list in Component **"Light key in Bottoms"** (e.g. n-Butane) and enter its value under Mole fraction label. Select Heavy Key component name from the list in Component **"Heavy Key in Distillate"** (e.g. i-Pentane) and enter its corresponding value under Mole fraction label. Enter Condenser and Reboiler Pressure (*NB: The value of Reboiler pressure is greater than the value of Condenser pressure*). Enter the value of External Reflux Ratio. This gives a highlighted green comment "**OK**" as shown in Figure K-20.

Figure K-20.

Snapshot of Column window with required Parameters.

Step 21. The PFD window shows the results of shortcut simulation with **Feed, Distillate** and **Bottoms** and their corresponding values of **Temperature, Pressure** and **Molar flow rate** as shown in Figure K-21.

Figure K-21.

Snapshot of PFD window showing the shortcut method simulation results.

Step 22. A summary of the printout of the data sheet can be obtained by right clicking on the column icon in Figure K-22 to display a menu with **Print Datasheet** as shown in Figure K-22. Then click on the **Print Datasheet** menu.

Figure K-22.

Snapshot of PFD window showing the **Print Datasheet** menu.

Step 23. Click on **Print Datasheet** to display “Select Data block(s) to Print for Shortcut Column” window as shown in Figure K-23. Click on **Preview** button to view the Shortcut Column datasheet results or alternatively on **Print** button to print the hard copy of the results.

Figure K-23.

Snapshot of **“Select Data block(s) to Print for Shortcut Column”** window.

Step 24. Display results of the Shortcut Column simulation.

Figure K-24.

Snapshot of shortcut Column simulation from Honeywell UniSim® software.

Appendix L. Simulation Results Using UniSim

Simulation Results Using UniSim® (Honeywell) Design Software

1. Simulation Results of Fractionation of Crude Distillation Unit.

2. Simulation Results of Gas Processing

3. Simulation Results of Reactive Distillation of Methyl Acetate Production.

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