Equilibrium Separation Column

In equilibrium separation processes, two or more coexisting zones are created with preferential distribution of the different components involved in the process in each zone. For example, in distillation a liquid and vapor zones are created and the components are separated in different proportions between these zones. In absorption and stripping, liquid and gas phases are created. In extraction, two liquid phases are created. The zones are assumed to be, as the name implies, in equilibrium with each other. This means that the temperature, pressure, and phase compositions are all in thermodynamic equilibrium. Thus, while all separation processes are essentially a mass transfer process, the equilibrium assumption cancels out the need for dealing with transfer *rates* and focuses only on the transfer *amounts*.

Equilibrium separation processes are usually operated in a counter current configuration in which the two zones are made to flow opposite to each other in a closed vessel (or column). To ensure good contact, the column is either equipped with trays or filled with packing. In both cases, the contact process can be viewed as a stagewise process in which the two zones are contacted thoroughly in each stage and leaves at equilibrium. Each zone then proceeds to a different stage where it is contacted again to leave at equilibrium, and so on. The counter current configuration provides better driving force for transfer than co-current configuration.

A schematic representation of typical (general) column used for equilibrium separation is shown in Figure 74. In this diagram, the material to be separated (F) is fed to the column (more than one feed can be employed). Depending on the feed state, it will either start to rise towards the top of the column or fall towards the bottom.

At the top of the column, a condenser can be used to condense part or all of the vapors that leave the top. This is called the overhead condenser. Part (and sometimes all) of the condensed liquids are returned to the column as a reflux. The reflux then combines with any liquid feed at the top of the column and flows down. The remaining part of the condensate is taken as a distillate or overhead product.

Liquids flow down the column, contacting the vapors as they rise, to the bottom where it can be partially vaporized in the reboiler. The reboiler heats the fluids and boils up the
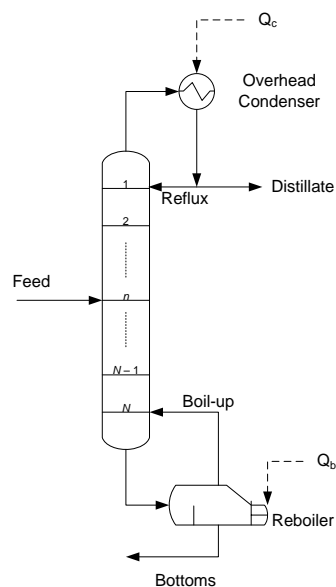
Figure 74. Simple, continuous distillation column

liquid which partially evaporates. The vapors (or boil-up) are returned to the column, and a bottom product, or bottoms, are drawn as a liquid.

As mentioned earlier, such separation processes are treated as an equilibrium system. With this assumption, it is possible to treat a distillation column using thermodynamic phase equilibrium relations. The validity of this assumption depends on the ability of each stage to achieve temperature, pressure, and phase equilibrium. Practical operating considerations impose a limitation on how close equilibrium is approached. Therefore, a tray efficiency is used to express the deviation from equilibrium. Using the tray efficiency allows the definition of theoretical and actual stages. A theoretical stage is one that achieves equilibrium between the phases, while actual stage represents an actual tray in the column.

While the application of the equilibrium approach and efficiencies had been successfully applied to many separation systems, a new approach has been developed that takes into account the transfer process between the phases. This approach is referred to as non-equilibrium (NEQ) or rate-based approach. In this approach, a stage (or tray) can be imagined to be divided into two phases, and material balances are written for the components in each phase. The rate dependence in the material balance arises from the mass transfer of the component from one phase to
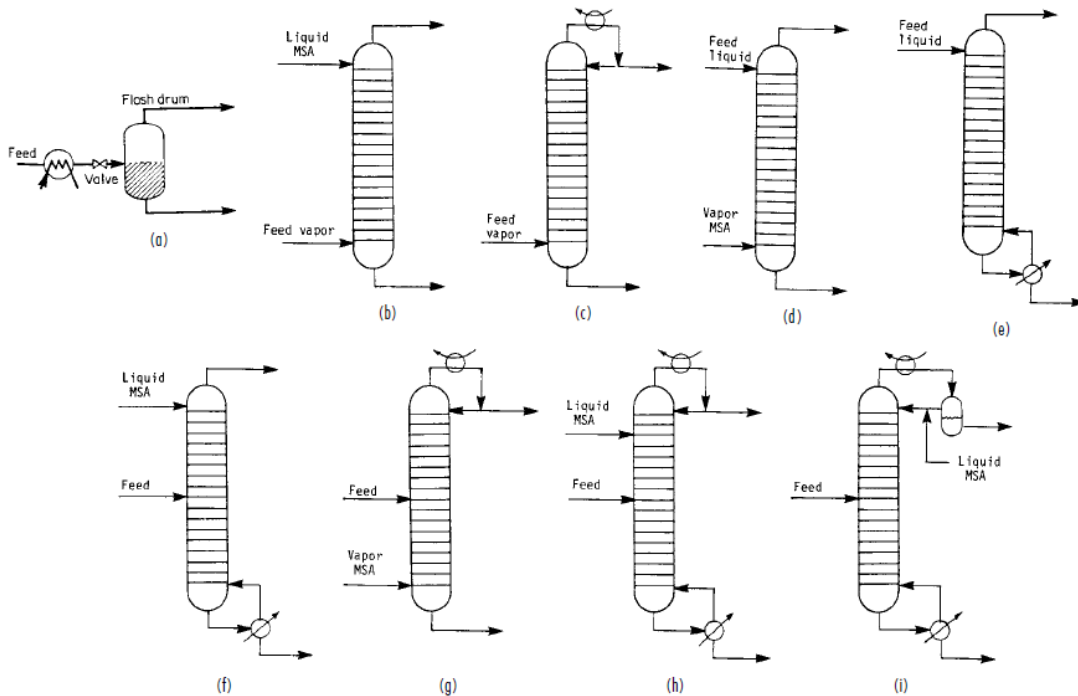

Figure 75. Separation operations related to distillation. (*a*) Flash vaporization or partial condensation. (*b*) Absorption. (*c*) Rectifier. (*d*) Stripping. (*e*) Reboiled stripping. (*f*) Reboiled absorption. (*g*) Refluxed stripping. (*h*) Extractive distillation. (*i*) Azeotropic distillation.[20]

another.

The above discussion and the configuration shown in Figure 74 present a distillation process in which the separation is based on the relative volatility of the components. Other processes related to distillation are shown in Figure 75. In addition to the phases present inside the column, variations in these processes involves differences in the feed(s), presence or absence of reboiler or condenser, and the use of mass separating agent (MSA). For example, in an absorption column a liquid stream is fed at the top of the column while a gas stream is fed from the bottom. The two phases are contacted in the column before leaving to further processing. No reboiler or condenser are used.

Operating Conditions

Multistage separation towers can be operated anywhere the two phases coexist, avoiding the proximity of the critical point. Pressure ranges for operation vary from very low (< 1 psia for vacuum distillation used with temperature sensitive materials) to several hundred (as is the case for absorbers operation).

The selection of operating pressure for distillation columns can be roughly estimated based on the distillate bubble point pressure as illustrated in Figure 76. Typical operating pressures in distillation range from 1 to 415 psia. The algorithm suggested in the figure starts by a known or assumed distillate composition and calculates the bubble-point pressure at ~50°C. Based on this pressure, the type of condenser is selected and the expected column operating pressure is designed based on a typical pressure drop inside the column of 5−10 psi. The temperature of the reboiler is then determined based on dew-point calculations taking into account any limiting temperatures that may be encountered (decomposition, critical, polymerization).

Distillation Column Design

The complete design of a distillation column requires the determination many parameters such as the feed location, number of trays or column diameter, trays design or packing type, separation between trays or packing height, total column height, and mechanical design. The complete design procedure for columns require the application of rigorous calculations, general rule of thumbs, and previous experience. One of the first design steps required for distillation column is to determine the number of theoretical stages required and the feed location.

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[20] Don Green and Robert Perry, *Perry's Chemical Engineers' Handbook, Eighth Edition*, 8th ed. (McGraw-Hill Professional, 2007).

The most rigorous analysis of distillation (and other staged separation processes) can be made based on a stage-by-stage analysis. This is done by performing components and energy balances on each stage assuming equilibrium is achieved between the two phases. Consider for example the stage (or tray) shown to the right (take the light phase to be vapor and the heavy phase to be liquid). In this stage, the liquid enters the stage j with a flow rate of L_{j-1} and composition $x_{i,j-1}$

and leaves with a flow rate and composition of L_j and $x_{j,i}$. A similar convention is used for the vapor with V and y. A fresh feed can be added to the stage with flow rate of F_j and composition
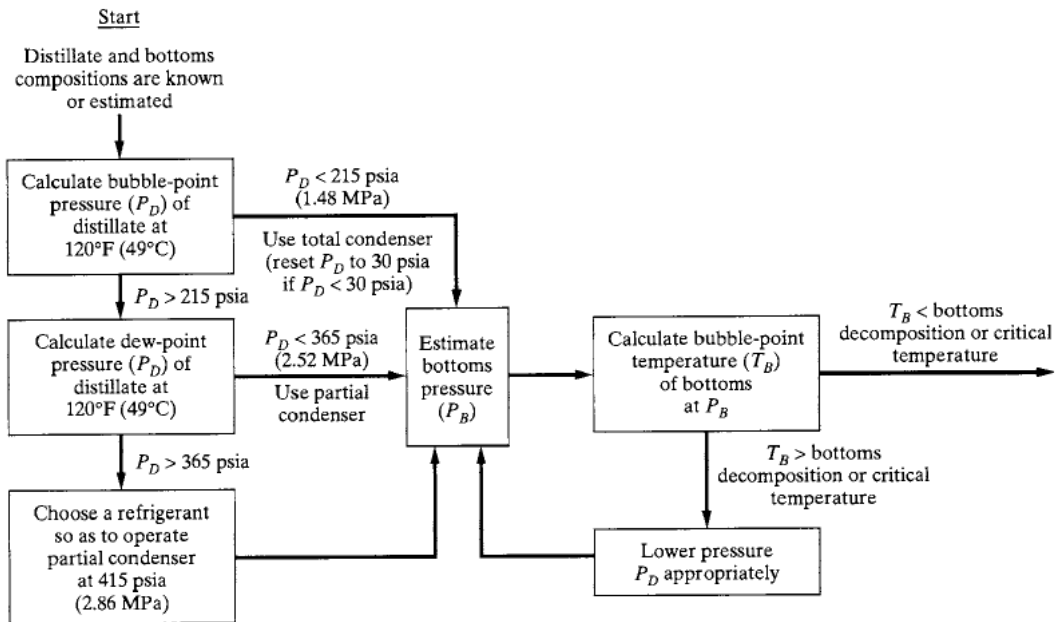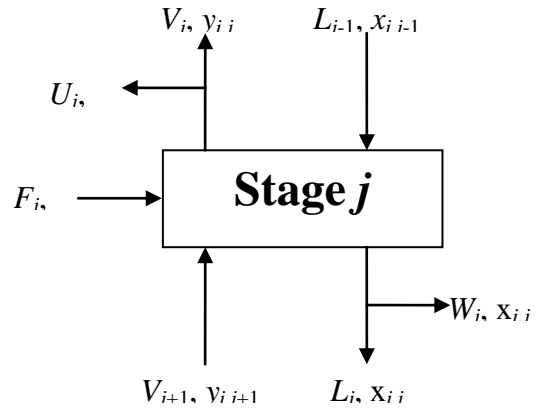

Figure 76. Preliminary selection of operating pressure for distillation columns.[21]

$z_{j,i}$. Side streams (W_j and U_j) can be drawn from the vapor and/or liquid streams. The total material balance on this stage is given by:

(48) $V_{j+1} + L_{j-1} + F_j - \left(1 + r_j^V\right)V_j - \left(1 + r_j^L\right)L_j = 0$

where $r_j^V = U_j/V_j$ and $r_j^L = W_j/L_j$ are the draw ratios. Similarly, a component balance can be written:

(49) $V_{j+1}y_{i,j+1} + L_{j-1}x_{i,j-1} + F_j z_i - \left(1 + r_j^V\right)V_j y_{i,j} - \left(1 + r_j^L\right)L_j x_{i,j} = 0$

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[21] Warren D. Seider, J. D. Seader, and Daniel R. Lewin, *Product and Process Design Principles: Synthesis, Analysis, and Evaluation*, 2nd ed. (Wiley, 2003).

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In addition, the heat balance of the stage is given by:

$$(50) \qquad V_{j+1} H_{j+1}^V + L_{j-1}H_{j-1}^L + F_j H_j^F - \left(1 + r_j^V\right)V_j H_j^V - \left(1 + r_j^L\right)L_j H_j^L - Q_j = 0$$

where H denotes the enthalpy.

The above model is completed by adding the equilibrium constraint on the products leaving the stage. This is usually expresses as:

$$(51) \qquad y_{i,j} = K_{i,j}x_{i,j}$$

where $K_{i,j}$ are the equilibrium ratios or K value for species i on stage j and is given by:

$$(52) \qquad K_{i,j} = \frac{y_i}{x_i} = \frac{\gamma_i f_i}{\hat{\phi}_i P}$$

The degrees-of-freedom available for the entire column can be obtained by analyzing a system of N stages. Take, for example, the simplest case of an absorption column (show in Figure 75(b)) with N stages. In this column if both feeds have known compositions and flow rates, we have $N(2c + 1)$ equations and $N(2c + 1)$ variables (for each stage). Thus, the degrees-of-freedom for this system is zero. If an equilibrium reboiler stage is used, the heat input to the reboiler becomes an additional degree of freedom is added. If a condenser is added, the degrees-of-freedom becomes two. If a side stream is drawn from any stage, the degrees-of-freedom becomes three, and so on.

The above model gives rigorous analysis of the separation column and detailed results of its behavior. However, the model requires the input of important design variables such as the number of stages, the feed location, in addition to the specifications of the feed and required products. In addition, the numerical analysis required to solve the model equations can be complex especially with multiple component systems and large number of stages. Therefore, simple (or shortcut methods) have been developed to address the above two points.

One of the simplest methods that was introduced for determining the number of stages and feed location was the McCabe-Thiele method. This method relies on representing the material balances (Equation (49)) and the equilibrium relation (Equation (51)) graphically and counting the number of theoretical stages and feed location based on simplifying assumptions. An example of the McCabe-Thiele is shown in Figure 77. The method divides the column into a top (or enriching) section and bottom (or stripping) section, drawing the overall material balance (operating lines) equation for each section, and counting the number of stages as shown in the figure. The feed location is found at the intersection of the operating lines (with the q-line). Refer to your separation textbook for the definitions and equations used to generate this diagram. One important notation used to plot the McCabe Thiele chart is the reflux ratio. This ratio is defined as the amount of liquid returned to the column (L) divided by the amount of overhead vapor (V), in other words:

$$(53) \qquad R = \frac{L}{V}$$

The reflux ratio ranges from a minimum (R_{min}, when the operating line intersects the equilibrium line) to unity (when the entire overhead vapor is refluxed).

Other shortcut methods have also been devised to provide quick estimation for distillation columns calculations. One of the methods is the Fenske-Underwood-Gilliland (FUG) shortcut method. This method is applicable to a single feed distillation column with only distillate and bottom products (no side draws), and gives the reflux ratio, number of equilibrium stages, and the location of the feed. The procedure for this method can be roughly summarized as:

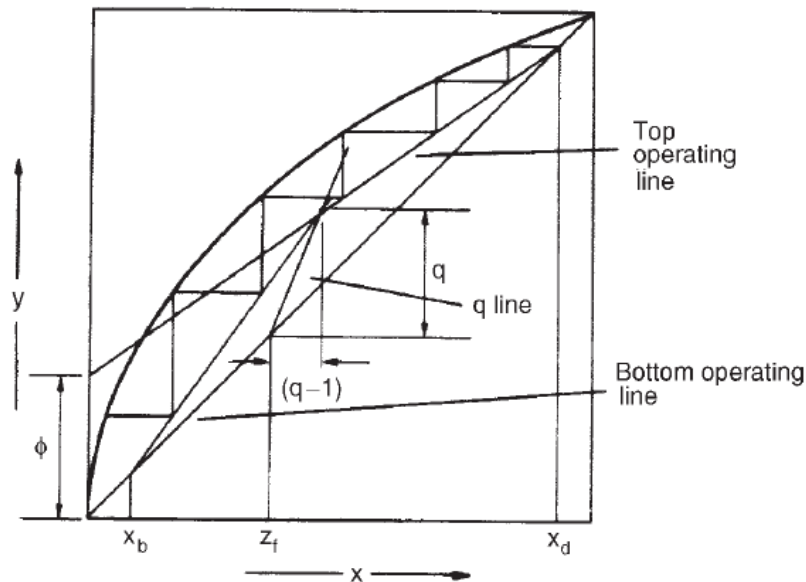1. Estimate the minimum number of stages at total reflux using Fenske's equation:

Figure 77. McCabe-Thiele diagram.[22]

$$(54) \qquad N_{min} = \log\frac{\left[\left(\frac{d_{LK}}{b_{LK}}\right)\left(\frac{b_{HK}}{d_{HK}}\right)\right]}{\log(\alpha_{LK,HK})}$$

where d and b are the flow rates in the distillate and bottoms of the light (LK) and heavy (HK), and $\alpha_{LK,HK}$ is the relative volatility of the light to heavy component and is equal to K_{LK}/K_{HK}, taken as constant value throughout the column.

2. Estimate the minimum reflux (R_{min}) from the solution of Underwood's equations:

[22] R. K. Sinnott, *Coulson & Richardson's Chemical engineering Design*, vol. 6, 4th ed. (Butterworth-Heinemann, 2005).

$$(55) \qquad \sum \frac{\alpha_i (x_{iD})_{min}}{\alpha_i - \Theta} = R_{min} + 1$$

where Θ is found from the solution of the following equation:

$$(56) \qquad \sum \frac{\alpha_i z_{iF}}{\alpha_i - \Theta} = 1 - q$$

3. Estimate the actual number of equilibrium stages for a specified reflux ratio using Gilliland correlation:

$$(57) \qquad \frac{N - N_{min}}{N + 1} = 1 - \exp\left(\frac{1 + 54.4\Psi}{11 + 117.2\Psi} \times \frac{\Psi - 1}{\Psi^{0.5}}\right)$$

with $\Psi = (R - R_{min})/(R + 1)$. A recommended choice for R is $1.1 - 1.5R_{min}$. Usually, a lower reflux is used for difficult separations while a higher value is used for easy separations.

4. Using Fenske's equation to calculate the minimum number of stages required for the enriching section. This is done by applying Equation (54) with the feed composition as the bottoms.

The method was modified by Winn to provide improvement over Fenske's equation, and the method is sometimes referred to as Winn-Underwood-Gilliland equation.

Column and Tray Efficiencies

Actual columns require more stages than theoretical ones. The number of actual stages needed can be related to the theoretical stages through either column or tray efficiency. The column efficiency (E_o) is defined as:

$$(58) \qquad E_o = \frac{N}{N_A}$$

where N is the number of theoretical stages and N_A is the actual number. The column efficiency can be affected by a number of factors and have been correlated to different properties. An example is shown in Figure 78 where the efficiency is correlated to the product of the relative volatility and liquid viscosity (the O'Connell correlation).
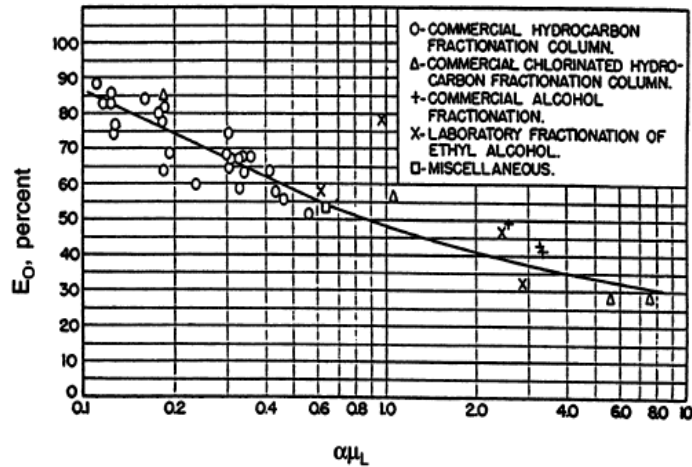
Figure 78. Correlation of column efficiency to $\alpha\mu_L$.[23]

The efficiency is also defined on tray basis. A common definition used for this purpose is the Murphree efficiency, defined for component i stage j as:

$$(59) \qquad E_{mV} = \frac{y_{i,j} - y_{i,j-1}}{K_{i,j}x_{i,j} - y_{i,j+1}}$$

where y_e is the equilibrium composition of the vapor phase leaving the stage. Another definition for the efficiency is the vaporization efficiency defined as:

$$(60) \qquad E_V = \frac{y_{j,i}/x_{j,i}}{K_{i,j}}$$

For packed columns, tray efficiencies are defined based on the height equivalent to a theoretical plate (HETP). The Murphree efficiency can be estimated from correlations such as the van Winkle's correlation:

$$(61) \qquad E_{mV} = 0.07 Dg^{0.14} Sc^{0.25} Re^{0.08}$$

where Dg is the surface tension number ($= \sigma_L/\mu_L u_v$).[24]

Tray Design and Operation

The internals of the column are designed to provide good contact between the phases. Trays are usually preferred for several reasons, but packing can be used especially with smaller column diameters. The simplest type of contact plates is the sieve plate which is made of perforated metal sheets. The plate is installed in the column as shown in Figure 79. Other types of trays, such as the bubble cap and valve trays are also used.

[23] Fouad M. Khoury, *Multistage Separation Processes*, 3rd ed. (CRC Press, 2004).
[24] Sinnott, *Coulson & Richardson's Chemical engineering Design*,6:.

Trays are designed to have either single or multiple passes. The number of passes represents the number of passages available on each tray. The passes split the liquid and force it to travel only
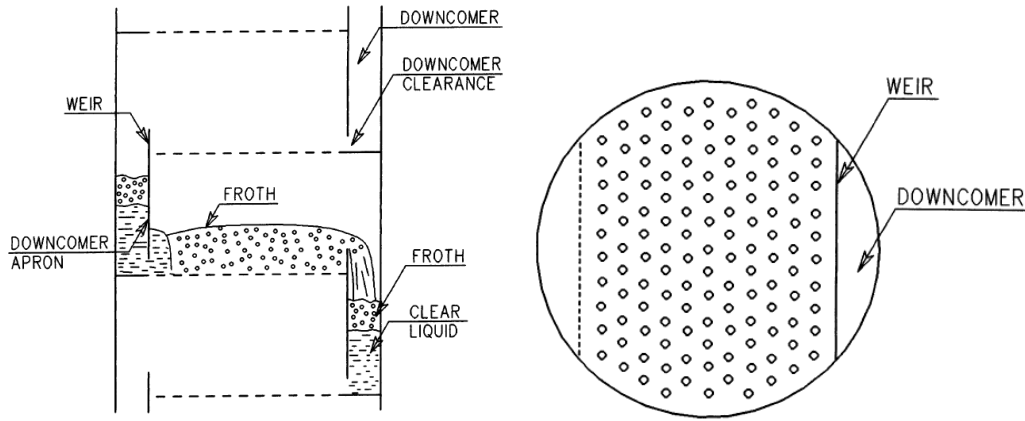

Figure 79. Sieve plate common terminology.[25]

certain portion of the tray.

The complete design of trays depends on the hydraulic analysis of the system, where the vapor and liquid flows around the tray are analyzed. This analysis includes the following phenomena:[26]
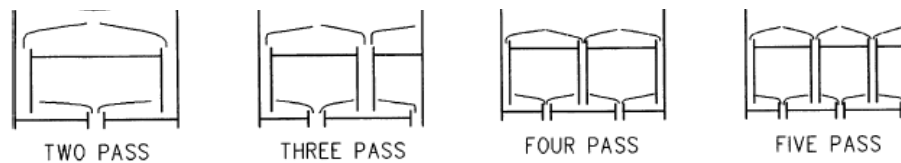

Figure 80. Single and multiple pass trays.

Foaming: The formation of froth on the trays is desirable for maximizing the interfacial area and mass transfer between liquid and vapor. Froth is formed through agitation and turbulence brought about by the vapor flow through the liquid. The amount of froth formed (foamability) and its tendency to linger (foam stability) are related to the physical properties of the liquid. High foamability can limit the allowable vapor flow, and high foam stability can limit the allowable liquid flow. It is clear that although the formation of foam is desirable, excessive foamability or foam stability should be avoided. The foaming properties must therefore be taken into account in the design and rating of column trays. In certain situations foamability may be adjusted by the addition of surfactants. The effect of foaming on the tray capacity is presented using a foaming factor. The factor values range from 1.0 (non-foaming systems) to less than 0.6 (highly foaming systems).

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[25] Khoury, *Multistage Separation Processes*.
[26] Taken from: Ibid.

Vapor Entrainment: As liquid flows down through the downcomer, some froth flows with it and some more is formed in the downcomer by the turbulent flow. The downflowing froth should be allowed to break up in the downcomer before it reaches the tray below. Any froth flowing to the tray below carries with it some vapor, resulting in vapor entrainment. This lowers the tray efficiency on account of vapor flowing in the wrong direction. The downcomer volume must be large enough that the residence time is sufficient to allow the froth to disintegrate.

Liquid Entrainment: When vapor disengages from the froth on a tray, liquid droplets may be entrained with it to the tray above. Liquid entrainment could be especially severe with highly foamable liquids where the froth occupies a large portion of the space between the trays. If the top of the froth is close to the tray above, the liquid droplets do not have enough time to fall back to the froth. Foamability therefore limits the vapor flow capacity of the trays. Handling high vapor flows with a foaming liquid requires larger tray spacing.

Tray Pressure Drop: The flow of vapor upwards in the column from tray to tray requires a column pressure gradient to overcome the pressure drops through the trays. There is a friction pressure drop resulting from the flow of vapor through the tray openings. This pressure drop would occur even with no liquid on the tray, and is appropriately referred to as dry tray pressure drop. The vapor exiting the tray openings finds itself at the bottom of the liquid holdup on that tray, where the pressure is higher than that in the vapor space above, due to the liquid head. Additionally, the pressure inside the vapor bubbles coming out of the tray openings is slightly higher than the liquid pressure due to surface tension. All in all, the total pressure drop between two trays, or the difference between the pressure in the vapor space on tray $j + 1$ minus the pressure in the vapor sure in the vapor space on tray j is the sum of three pressure drops. In terms of clear liquid heads:

(62) $\qquad h_t = h_d + h_l + h_\sigma$

Each term in the above equation is calculated from fluid mechanics principles or correlations to estimate the total pressure drop.

Weeping: The excessive flow of liquid through the tray openings at low vapor rates and high liquid rates is a pressure-related phenomenon. If the liquid head on the tray exceeds the sum of the dry tray pressure drop and the surface tension pressure drop weeping can be expected. It can be prevented during the design phase and/or during operation by ensuring that the vapor and liquid flows satisfy the condition that $h_d + h_\sigma > h_l$.

Liquid Holdup: The volumetric tray liquid holdup can be considered consisting of two main parts: the liquid on the active tray area and the liquid in the downcomer. These holdups can be estimated from the liquid heights and the corresponding areas.

Flooding: When liquid accumulate on the tray, they buildup pressure than can hinder vapor flow through the column. When the holdup increases, the vapor upflow can be affect to an extent that

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will severely reduce the tray efficiency and increases the pressure drop significantly inside the column.

Figure 81 shows some typical operating problems with sieve trays column and its relation to the vapor and liquid flows.
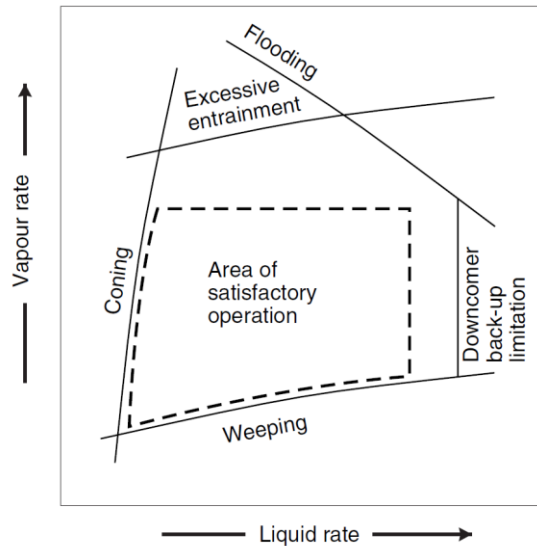

Figure 81. Sieve plate performance diagram.[27]

Separation Column in Aspen Plus

Aspen Plus provides several modules for multistage separation processes. These models are the DSTWU, Distl, and RadFrac (which are shortcut and rigorous methods for distillation, absorption, and stripping application), Extract (for extraction applications), MultiFrac, SCFrac, and PetroFrac (for advance fractionation applications), and BatchSep (for batch distillation).

DSTWU Module
This module provides a shortcut for distillation column with single feed and only distillate and bottoms products using the Winn-Underwood-Gilliland methods. The method is useful for making initial estimates on the stage number and feed stage for the process, as well as providing initial estimates for the rigorous methods.

The input form for the DSTWU block is shown in Figure 82. The required information is simply the number of stages or reflux ratio, the light and key components, the pressure profile, and the type of condenser. The block can perform a case study of the number of theoretical plates versus the reflux ratio, and perform packed columns calculations of HETP. Both these options are available from the **Input | Calculations Options** form.

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[27] Sinnott, *Coulson & Richardson's Chemical engineering Design*,6:.

Figure 82. Input form for the DSTWU block.

As an example, consider a stream of 40 propane:60 isobutane at 322 K and 20 atm, and take a basis of 1 kmol/sec. The stream is to be separated with 99% propane and 0.1% isobutene recoveries in the distillate. Setup a simulation with the above information and an R/R_{min} of 1.15, assuming 19 atm and 20 atm condenser and reboiler pressures, and check out the Results form. You will see that the program has calculated the minimum and actual reflux ratios, the minimum and actual number of stages (43 stage), the feed stage (25), and the

reboiler and condenser temperatures and duties. You can also calculate the HETP for the column if the actual packing height is input. In addition, select the "Generate table of reflux…" and run the simulation again. If you check the Results folder again you will see that a case study of reflux versus number of stages have been plotted.

Figure 83. Input form for the Distl block.

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Distl

Similar to the DSTWU model, this model performs calculations for a single feed and only distillate and bottoms products. The input to the block is the number of stages, the feed stage, the reflux ratio, the distillate to feed flow ratio, and the pressure profile. The input window is shown in Figure 83.

If we repeat the previous example with the results as the input for this method, we find that the compositions of benzene in the distillate and bottoms streams are 99.99 mole% and 3.2×10^{-5} mole%. These figures are close to that obtained from the DSTWU method.

RadFrac

The RadFrac model is the main separation block in Aspen Plus. The block can perform simulation, sizing, and rating of tray and packed columns. The model specification, done in the Setup form, requires complete specifications of column configuration, specifications, feed(s), product(s), and any side streams. The number of specifications required depends on the degrees of freedom available and can be based on the distillate or bottom rates, the reflux or boilup rates, the reflux or boilup ratio, the condenser or heat duties, the distillate or bottoms ratio to the feed, or any combination of the above.

In addition, the pressure profile inside the column must be specified. This is done in the **Setup | Pressure** form where the pressure can be input as a constant, fixed pressure drop, or a pressure profile is given.

The condenser and reboiler can also be configured. Default configurations are automatically input once you select the exchanger types. For the condenser, you can specify its type (total or partial) and the degrees of subcooling. For the reboiler, two types are available, kettle and thermosiphon. No specifications are needed when a reboiler is used, but the thermosiphon requires determination of its configuration and one or two specifications.

Figure 84. Input form for the RadFrac block.

The RadFrac model allows connection of side product streams to any stage (up to three from each stage), plus other auxiliary side streams (called pseudo streams) that can be used to report data and does not affect simulation.

If you try to simulate the previous example using the RadFrac model, you will notice some differences in the results. For example, the distillate temperature in RadFrac is 302 K, which is close to that obtained with the Distl block but much lower than the DSTWU block (333 K). The reboiler with RadFrac has a temperature of 341 K which is lower than the other two blocks.

Two important features are available with the RadFrac model: Design Specs and tray and packing sizing and rating. The Design Specs feature allows manipulating the columns specifications to obtain required output. For example, if we want to obtain 99.9 mol% propane from the system in the previous example, we can setup a Design Spec for the mole purity of 0.999 of propane in the distillate stream, then setup a Vary variable based on one of the two specifications using the **Setup** form. For example, setup a Vary case for which the Distillate to feed ratio is the manipulated variable and set its range from 0.1 to 0.9. Then, the feed rate will be adjusted to obtain the required specification. In this case, the distillate to feed ratio is calculated as 0.397.

The sizing and rating capabilities can be used to design or size some aspects of the column internals. For example, if you want to use sieve trays for all the stages, you can add a new Tray Sizing case, select the Sieve tray and number of passes for all stage. Once you run the simulation, the tray diameter, column diameter, down comer area, and weir height will all be calculated. The rating calculations will give some operating parameters such as the flooding factor, maximum backup in the downcomer area, and pressure drop.

> **NOTE:** Columns, especially distillation columns, are among the more difficult units to setup and can cause convergence problems in the simulation. This is due to the many iterative calculations needed inside the column. In case the simulation fails to converge, there are some typical points to check. The following are suggestions from Aspen taken from the same source of this exercise:
>
> 1. Check that physical property issues (choice of Property Method, parameter availability, etc.) are properly addressed.
> 2. Ensure that column operating conditions are feasible.
> 3. If the column err/tol is decreasing fairly consistently, increase the maximum iterations on the RadFrac Convergence Basic sheet.
> 4. Provide temperature estimates for some stages in the column using the RadFrac Estimates Temperature sheet (useful for absorbers).
> 5. Provide composition estimates for some stages in the column using the RadFrac Estimates Liquid Composition and Vapor Composition sheet (useful for highly non-ideal systems).

6. Experiment with different convergence methods on the RadFrac Setup Configuration sheet.

>> When a column does not converge, it is usually beneficial to Reinitialize after making changes.

Exercise 1: DSTWU

A mixture of 50 mol/hr of benzene, 50 mol/hr of toluene, and 50 mol/hr of paraxylene is to be distilled at 1 atmosphere. The feed is a saturated liquid at 1 atmosphere. The column is to operate at a reflux ratio of 2.8, and a 1 psi pressure drop across the column is anticipated. Ninety five percent of the benzene fed and no more than 5% of the toluene fed is to be taken overhead. Use the IDEAL property method and the shortcut distillation block DSTWU to estimate the column performance. Create the flowsheet as shown below.
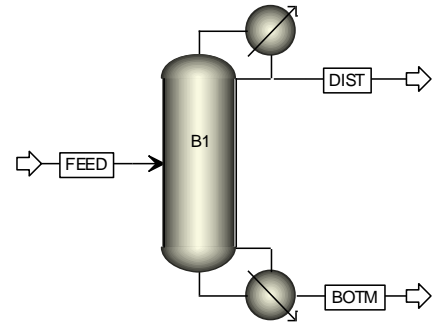
Questions:

15. What are the outlet temperatures from the column?

    Distillate:............., Bottoms:...

16. What is the heat duty for the condenser and reboiler?

    Condenser:............., Reboiler:..

17. Fill out the following table:

Benzene Recovery (%)	Number of trays
95	....
99.5	....
99.95	....

18. Prepare a plot of reflux versus theoretical stages.

19. Repeat Question 2 using the NRTL model.

    Condenser:............., Reboiler:..

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Exercise 2: Dstil and RadFrac

Repeat Exercise 1 using the Distl and RadFrac blocks. Use the results from Exercise 1 for the numb er of stages, location feed, and the distillate to feed ratio.

Questions:

1. Summarize your results in the table below:

	DSTWU	Distl	RadFrac
Distillate T (^{o}C)	....	....	....
Bottoms T (^{o}C)	....	....	....
Condenser Duty (kW)	....	....	....
Reboiler Duty (kW)	....	....	....
Benzene in distillate (mol%)	....	....	....
Benzene in bottoms (mol%)	....	....	....

Exercise 3: Detailed Distillation Column Design

Simulate the system shown in the diagram below.

FEED

Feed:
63.2 wt% Water
36.8 wt% Methanol
Total flow = 120,000 lb/hr
Pressure 18 psia
Saturated liquid

COLUMN

DIST

Column specification:
38 trays (40 stages)
Feed tray = 23 (stage 24)
Total condenser
Top stage pressure = 16.1 psia
Pressure drop per stage = 0.1 psi
Distillate flowrate = 1245 lbmol/hr
Molar reflux ratio = 1.3

BTMS

Use the NTRL-RK property method
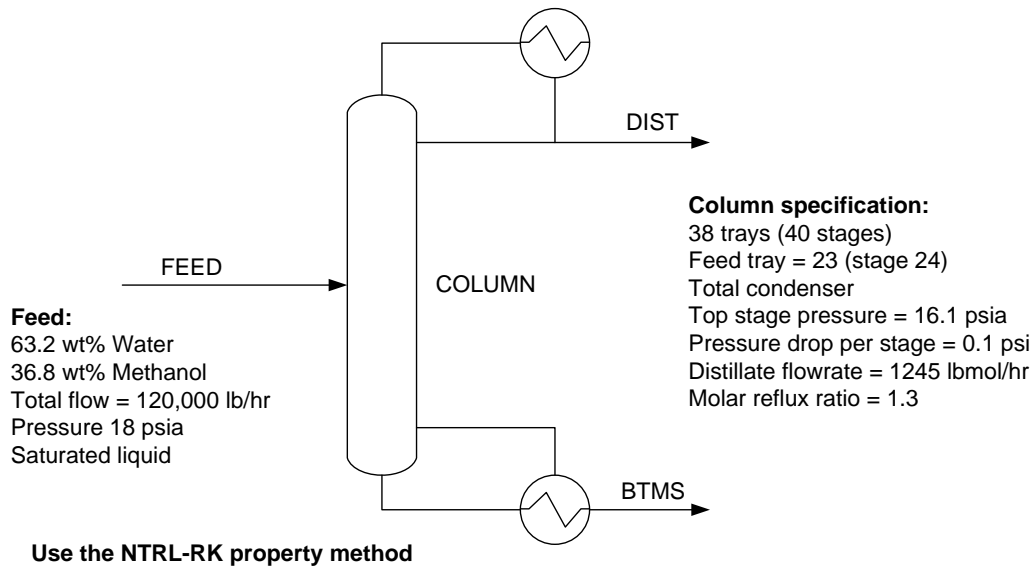
Questions:

1. Generate a temperature profile plot inside the column. Also, generate a plot showing the water and methanol composition in the column

2. If the column trays have a Murphree efficiency of 65%, and the reboiler and condenser have an efficiency of 90%, what will be the heat duty for the reboiler and condenser?

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3. Experiment with the sizing form for the column and determine the tray diameters if a bubble-cap trays are used. What is the required column diameter? What is the range for the tray diameters?

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4. If we have a 8' diameter, bubble cap trays with 6" weir height and 3" cap diameter. Perform tray rating and determine the flooding factor and the pressure drop per tray.

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