

Reactions and Separations

Shortcut Distillation Calculations via Spreadsheets

JAKE JEVRIC,

PPG CHEMFIL

MUHAMMAD E. FAYED,

RYERSON UNIVERSITY

This method uses a numerical solution to a McCabe-Thiele diagram to find the theoretical number of stages for binary and pseudo-binary systems, then calculates the actual number of stages, reflux ratio and column dimensions.

It is often necessary to develop data for a range of operating conditions, so that the optimum configuration of a distillation tower can be found. There are two conventional methods to perform such a task, either graphically by hand (which is somewhat inaccurate and time-consuming) or by any of a number of commercial simulations that are faster, but costly to license. A third alternative is presented here: Merging the graphical and manual computational methods so that the inaccuracies of the former are compensated for by the speed of the computations. The calculations can be run by any spreadsheet program, such as Microsoft Excel, eliminating the need for employing expensive simulation software and for laboring over hand calculations. Further, the time involved from a programmer's point of view is no more (or considerably less) than that required to learn how to use a commercial simulation package.

The method presented here is easy to learn, and offers a quick way to make preliminary estimates of the tower diameter and height, number of stages, energy consumption, and reflux ratio. Although the calculation procedure is intended for binary systems, ternary systems can also be modeled if the third component is less than 10% by volume and its volatility is not drastically different from the those of the remaining two components.

Spreadsheet calculation procedure

The first step in any binary distillation calcula-

tion (apart from performing the mass balance around the column), is determining the vapor/liquid equilibrium (VLE) data. Raoult's law is used to calculate the saturation pressure for the pure components. Since most systems are non-ideal, the Van Laar equation is then applied to determine the liquid and vapor compositions. This equation includes the activity coefficients for a mixture, making it suitable for non-ideal systems ((1), p. 32). An ethanol/water system is used to illustrate the overall method.

Apply Raoult's law and the Van Laar equation

Use Raoult's law to find the saturated vapor pressure for each component. Published data are available for various compounds ((2), p. 10-141). The saturated vapor pressure of each component is expressed as:

$$P_{1sat} = 10^{[A - B/(T + C)]} \quad (1)$$

where P_{1sat} is the saturated vapor pressure of component 1 (mm Hg), T is the temperature ($^{\circ}\text{C}$), and A , B and C are the Raoult's law constants for each compound. For ethanol, $A = 8.32109$, $B = 1,718.1$ and $C = 237.52$. The values for water are: $A = 8.07131$, $B = 1,730.63$ and $C = 233.42$. To calculate the saturation pressure of a component, simply substitute the value of the temperature. In these calculations, 1 refers to ethanol and 2 is for water.

The liquid and vapor mole fractions of ethanol (x_1 and y_1 , respectively) are found using the total system pressure, P_{total} :

$$x_1 = P_{1sat} / (P_{1sat} + P_{2sat}) \quad (2)$$

$$y_1 = (x_1 P_{1sat}) / P_{total} \quad (3)$$

For an ideal system, these equations will determine the VLE data needed to estimate the column dimensions. However, since most systems are non-ideal, the Van Laar equation is applied. This equation calculates the partial pressures, given the mole fraction and saturation pressure of each component ((1), p. 44):

$$P_1 = x_1 P_{1sat} \exp \left(A_1 \left(\frac{A_2 - x_1 A_2}{A_1 x_1 + A_2 - x_1 A_2} \right)^2 \right) \quad (4)$$

$$P_2 = (1 - x_1) P_{2sat} \exp \left(A_2 \left(\frac{A_1 x_1}{A_1 x_1 + A_2 - x_1 A_2} \right)^2 \right) \quad (5)$$

where the exponential terms are also known as the activity coefficients, γ_1 and γ_2 , respectively. A trial-and-error approach is used to generate the VLE data. A

value of the liquid mole fraction of a component is chosen (here, for ethanol), and then a temperature is guessed that corresponds to that value of x_1 . The temperature is inputted into Eq. 1 to determine the saturation pressure of each component via Raoult's Law. Then the partial pressures are calculated using the Van Laar equation. Different temperatures are tried until the individual vapor pressures, P_{1sat} and P_{2sat} , yield values of P_1 and P_2 (the partial pressures) that add up to P_{total} . Values of A_1 and A_2 in Eqs. 4 and 5 for various two-component systems are available in the literature, e.g., in "Perry's Handbook" (3).

Often, it is easier to determine the temperature by simply finding when the sum of the partial pressures divided by the total pressure equals 1 (within a specified tolerance limit). Once this temperature is reached, the following equation is solved for y_1 , given the values for the other variables:

$$P_1 = y_1 P_{total} = x_1 \gamma_1 P_{1sat} \quad (6)$$

The mole fractions of the second component are simply determined:

$$y_2 = 1 - y_1 \quad (7)$$

$$x_2 = 1 - x_1 \quad (8)$$

A spreadsheet is constructed by linking its cells to the proper constants and equations. VLE data are needed at various temperatures and pressures in the column. To obtain these data points, the pull-down command is used, which incrementally updates each outsourced cell that is linked to an equation.

An advantage of this spreadsheet is that the total system pressure can be varied to account for a pressure drop through the column. To do this, an equation is written that reduces the total pressure by an increment for each cell until at the last cell, the final pressure is reached. Hence, by using this formula for the pressure at each interval of the VLE calculation, a pressure drop can be simulated using the Van Laar equation.

For instance, if the pressure at the top of the column is 760 torr, and 3,040 torr at the bottom, and 20 calculation cells are specified, then for stage n from the bottoms (B) of the column, the total pressure would be:

$$P_{total, n} = 3,040 - n(3,040 - 760)/20 \quad (9)$$

In general:

$$P_{total, n} = P_B - n(\Delta P) / (\text{No. VLE data points}) \quad (10)$$

A series of data points is needed so the program can fit the generated VLE data to so-called trendline curves

Nomenclature

$a, b, c, d, e, f, g,$	= trendline constants, dimensionless
A, B, C	= constants for Raoult's law, dimensionless
A_1, A_2	= constants in Van Laar equation, dimensionless
E_o	= stage efficiency
n	= number of data points
P	= pressure, mm Hg
q	= quality of liquid
T	= temperature, °C
x	= liquid mole fraction
y	= vapor mole fraction

Greek letters:

α	= relative volatility
γ	= activity coefficient
μ	= viscosity, N-s/m ²

Subscripts:

1, 2	= component 1 or 2
B	= bottoms
D	= distillate
F	= feed
i	= data point
l	= mixture
sat	= saturated
R	= reflux
VLE	= at vapor/liquid equilibrium

Table 1. Spreadsheet for the tabulation of VLE data.

Temp.	Partial Pressures, P_p , mm Hg		Vapor Pressures, mm Hg		Sum P_p = P_{total} ??	Ethanol Mole Fractions		Water Mole Fractions		Iteration
°C	Ethanol	Water	Ethanol	Water	Sum P_p/P_{total}	x_1	y_1	x_2	y_2	
NA	NA	NA	NA	NA	NA	0	0.0000	1	1.0000	0
140.98	232.691	2791.124	6051.404	2811.738	1.000	0.0075	0.0770	0.9925	0.9230	0.25
140.15	298.7444	2719.387	5913.987	2746.029	1.000	0.01	0.0990	0.99	0.9010	0.33
138.6	416.7961	2588.998	5664.127	2626.661	1.000	0.015	0.1387	0.985	0.8613	0.5
137.2	519.1505	2474.983	5445.861	2522.504	1.000	0.02	0.1734	0.98	0.8266	0.66
136.55	565.3373	2423.222	5346.867	2475.302	1.000	0.0225	0.1892	0.9775	0.8108	0.75
134.8	687.7966	2287.219	5087.579	2351.784	1.001	0.03	0.2315	0.97	0.7685	1
131.77	872.3044	2064.94	4662.822	2149.823	1.000	0.045	0.2971	0.955	0.7029	1.5
129.4	1005.28	1900.723	4351.088	2001.925	1.001	0.06	0.3464	0.94	0.6536	2
125.7	1172.278	1662.17	3898.513	1787.743	1.001	0.09	0.4138	0.91	0.5862	3
113.2	1347.797	1001.366	2644.298	1198.181	1.000	0.3	0.5738	0.7	0.4262	10
111.9	1334.727	944.8544	2535.623	1147.431	1.000	0.33	0.5854	0.67	0.4146	11
110.6	1318.559	890.3165	2430.653	1098.472	0.999	0.36	0.5964	0.64	0.4036	12
84.4	872.3808	95.95037	963.9394	422.7638	1.001	0.9	0.9019	0.1	0.0981	30
82.5	835.5953	64.2663	896.1136	392.0577	1.002	0.93	0.9303	0.07	0.0697	31
80.4	793.4814	34.81331	825.8483	360.3244	0.999	0.96	0.9570	0.04	0.0430	32
78.3	760.2665	0	760.2665	330.7832	1.000	1	1.0000	0	0.0000	33

over the full range of x and y values (0–1). These trendlines are actually polynomial equations, and are created for both $y = f(x)$, and $x = f(y)$. The number of data points is determined by observing how the trendline fits the curve. Excel automatically plots the calculated points. If there are not enough data points, the trendline will not fit the data correctly, but miss and skew, especially at the inflection points.

Once the parameters are found for the trendlines, an algorithm is used to apply the McCabe-Thiele method to determine the number of stages and the reflux ratio. This process is explained in detail later on.

Calculate VLE data

Table 1 presents the spreadsheet for the ethanol/water binary system (only the beginning, middle and ending sections are shown). The program also includes cells for the Van Laar coefficients and those for Raoult's law, which are not shown in Table 1.

The x_1 values are chosen by the programmer. The number of these values chosen determines, of course, the number of data points. All other entries in the table are either constants, or functions of constants and the inputted temperature.

To find the partial pressures of both components (Columns 2 and 3), the saturated vapor pressures are needed at a given temperature (Columns 4 and 5). The saturated vapor pressures are calculated using Raoult's law, and the partial pressures by applying the Van Laar equations. Since the total pressure is predetermined, a test column is set up in which the total pressure (the sum of the partial pressures) is divided by the actual total pressure at that stage (Column 6). When this value in Column 6 approaches unity (within a given tolerance) the correct temperature has been found.

Thus, once values of x_1 are chosen, the only variable is the temperature (Column 1). The last column shows the iteration number n in Eq. 10, which functions as a counter when a pressure drop is deemed necessary. In this example, the 33 in the last column is the number of VLE data points generated by the program.

Fractional calculations may be needed

The few fractional counters at the start of Table 1 were used to give further detail in areas that required it (for trendline accuracy). They represent additions that were implemented after the spreadsheet was fully assembled. There were insufficient data points, and the trendline did

Table 2. Trendline constants.

	$y = f(x)$		$x = f(y)$	
	Upper Section	Lower Section	Upper Section	Lower Section
a	0	-7616.9	0	0
b	-5.7695	8734.2	0	21.214
c	19.214	-4004	12.495	-20.508
d	-24.36	950.66	-36.041	7.6362
e	15.121	-128.77	35.812	-0.9711
f	-4.2069	10.862	-12.262	0.1403
g	1	0	1	0

Table 3. Trendline segmentation conditions.

Calculation Parameters for Segmented Trendline			
Lower segment y data range		Lower segment x data range	
0	0.5854	0	0.33
Upper segment y data range		Upper segment x data range	
0.5854	1	0.33	1

not fit the data set. Thus, a refining data set was added based on fractional pressure drops for the VLE data.

These fractional counters are a strong point of this method, in that if areas of the spreadsheet require additional data points, it is not necessary to construct a new spreadsheet. Instead, a simple insertion command within the spreadsheet suffices, and the same formulas can be reused.

Once the data points are calculated, a VLE diagram is plotted (Figure 1). Trendlines are needed of both $y = f(x)$, and $x = f(y)$ since both are used in modeling the McCabe-Thiele diagram. Due to the limitations of the trendline function in the spreadsheet used, the data were segmented into two parts for each curve. The curves were fitted to sixth-order polynomials (Tables 2 and 3).

The range for each polynomial ends at the inflection point of the x, y curve. However, the program found the ranges by trial and error.

If the data were not segmented, the trendlines would not produce an acceptable fit. The transition points

for the x and y VLE values are indicated below the equations in Figure 1. Once the graphs are constructed, the modeling process begins by displaying the respective trendline equations for each of the VLE curves and using them in the spreadsheet.

Determine the theoretical number of stages

The theoretical number of stages is calculated using the McCabe-Thiele method ((4), p. 402) (Figure

2). The following algorithm is used to generate the appropriate data points and "construct" the diagram (x_D is the distillate composition):

1. Start at the $y = x$ line for the distillate conditions (x, y) = (x_D, x_D).
2. Follow the constant y line to the VLE line point (x, y) = (x_{VLE}, x_D).
3. There are two options. For total reflux, follow the constant x_{VLE} line to the $y = x$ line for total reflux to the point (x, y) = (x_{VLE}, x_{VLE}). Alternatively, follow the x_{VLE} line to the minimum or optimum reflux line: $y = mx + b$ to the point (x, y) = ($x_{VLE}, m x_{VLE} + b$).
4. Repeat the algorithm for the new data point set defined in Step 3.

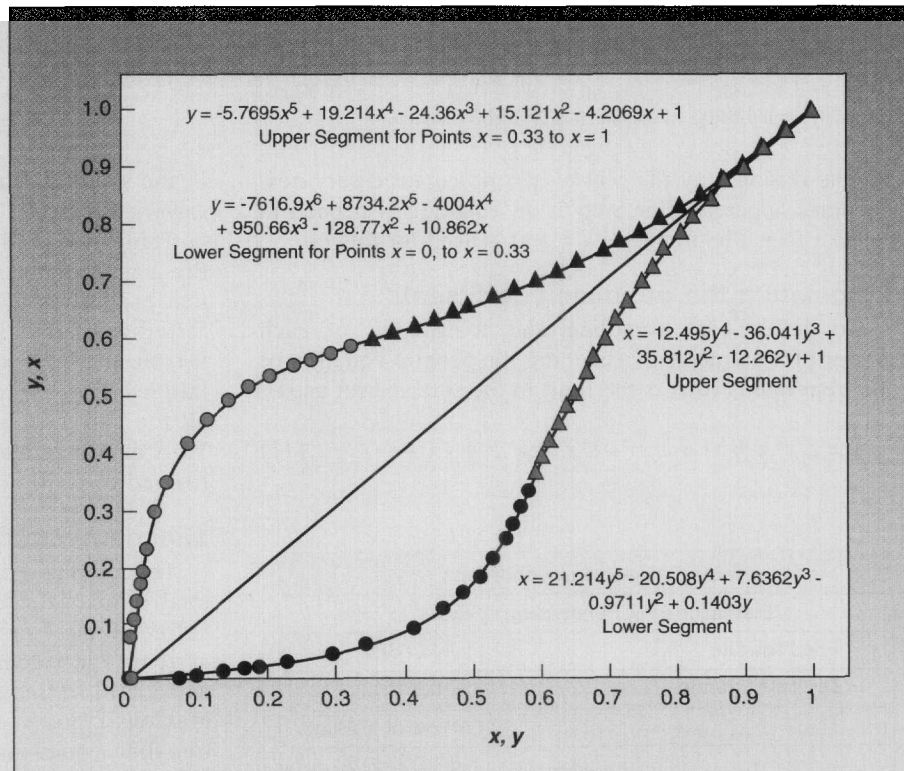


Figure 1. Trendlines were fitted to data points and are used in the McCabe-Thiele method.

Table 4. Tower operational conditions.

Distillation Parameters		Column stage spacing, in.	18
Parameter	Numeric Value		
$x_{Distillate}$	0.85	Column flooding velocity, fraction 0.95 If saturated, liquid slope = infinite = 1.00E+100	
$x_{Bottoms}$	0		
x_{Feed}	0.5		
m_q -line $R = \text{No.} \cdot R_{min}$	1.00E+100 1.5		

Eq. 11 represents the $y = f(x)$ trendline. For the $x = f(y)$ version:

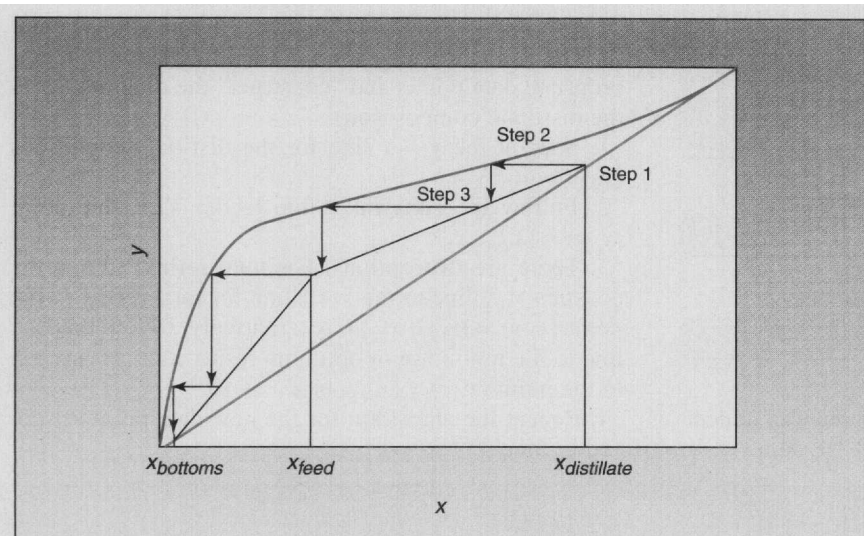
$$x = a y^6 + b y^5 + \dots + g \quad (12)$$

The values for the coefficients are unique for each line, i.e., the a in Eq. 11 does not have the same value as the a in Eq. 12. Tables 2–6 are set up to find the coefficients and perform the succeeding calculations.

Table 2 shows the results of the trendline calculations to find the coefficients of the polynomials. The conditions of the trendline segmentation (Table 3) determine where a conditional “if” statement will be used to jump from the upper to the lower lines within the spreadsheet calculation matrix.

Column specifications

Table 4 presents the feed, distillate and bottom specifications for the column, as well as other operating parameters. The slope of the operating line is taken as 10^{100} (1.00E+100) for a saturated liquid — this represents an infinite slope for calculation purposes. A reflux ratio of 1.5 was chosen arbitrarily and the column flooding-velocity fraction was set at 0.95, since this



■ Figure 2. McCabe-Thiele diagram for determining the number of stages in the rectifying and stripping sections of a distillation column.

The reason why the plot of $x = f(y)$ is necessary now becomes apparent. For Step 2, an equation is needed to relate x to y . The trendline $x = f(y)$ is used for this task.

Calculating the trendline coefficients

To define each trendline, the coefficients of each power of x are needed, assuming the generic equation of the trendline is (this is the limit in the spreadsheet used):

$$y = a x^6 + b x^5 + c x^4 + d x^3 + e x^2 + f x + g \quad (11)$$

is the general flooding velocity to which distillation towers are sized.

Table 5 calculates the flow conditions, based on the mass balance of the system. Table 6 lists the physical properties of the two components. Ethanol is referred to as the “graphed” component to differentiate it from the other component (water) used in the calculations. These three tables (Tables 4, 5 and 6) contain all the data needed to define the system, now that the VLE data are determined. Next, calculations are performed to find the minimum number of stages.

Minimum number of stages

The minimum number of ideal stages, which occurs at total reflux, is calculated using the algorithm presented before. Table 7 presents the results for the ethanol/water system. The values for the first row (x , y) are (x_D , y_D), and for all subsequent rows are (x_{VLE} , y_{VLE}), based on the previous x and y values, which are calculated using the trendline equations. Only three sections of Table 7 are shown due to space limitations.

This spreadsheet was set up to calculate compositions

Table 5. Tower flow conditions.

Distillation Flow Parameters, kg-mol/s	
Feed flowrate	0.06815
Distillate flowrate	0.038286111
GPM flow of distillate	
	31.67718152

for 20 stages. This number is usually more than sufficient to accommodate most distillation columns.

The calculations generated a value of $x = 0.002073297$ for Stage 11, which is a bottoms of essentially equal to zero.

Thus, 11 equilibrium stages are the chosen minimum number for this system. At Stage 12, x would be in the order of 3×10^{-4} , which is an unreasonable value.

Since the trendline is segmented, a conditional "if" statement is needed to discern when the calculations are within the ranges of the upper and lower trendlines. Thus, an "if" condition is put in the x cell calculation with the condition that "if" $y < y_{transition}$, then the operation is performed using the lower segment $x = f(y)$, otherwise the upper segment is used: $x = f(y)$. The $y_{transition}$ corresponds to the $x_{transition}$ value of 0.33 from Table 3, in this case, y has a value = 0.5854.

Table 6. Tower chemical species parameters.

Viscosity data, cP	Graphed Component	Non-graphed Component
	1.12	1
Latent heat, kJ/kg	842	2300
Specific heat, kJ/kg.°C	2.436	4.184
Condenser undercooling, °C		20
Surface tension, dyne/cm	22.80	73.40
Liquid density, lb/ft ³	49.32	62.45
Vapor density, lb/ft ³	0.089	0.035

Vapor density is usually about 0.15 kg-mol/m³ As such, convert from this.

Note: T unit = specific heat unit

1,000 N/m = dyne/cm

Molecular Weights

Graphed	46
Non-graphed	18

Slope of minimum reflux ratio operating line

Now the minimum reflux must be calculated. This is done by finding the intersection of the q -line (the quality of the liquid) with that of the operating line and the VLE curve. Based on the distillate, this point occurs at:

$$m_{q-line} = (y_{VLE} - y_F)/(x_{VLE} - x_F) \quad (13)$$

The subscript "F" refers to the feed. All of the above values are known except x_{VLE} , and y_{VLE} . Thus, y_{VLE} is expressed by way of the $x = f(y)$ trendline equation as:

$$m_{q-line} = (f(x_{VLE}) - y_F)/(x_{VLE} - x_F) \quad (14)$$

Since both the slope and the x_{VLE} point are unknown,

trial and error is used for simplicity to solve Eq. 14. The cells used for this calculation are shown in Table 8.

For a saturated liquid whose q -line is vertical, a very large slope is calculated (see the upper part of Table 8). The lower part of Table 8 shows the slope of the operating line for minimum reflux:

$$m_{Rmin} = (x_D - y_{VLE})/(x - x_{VLE}) \quad (15)$$

Table 7. Calculating the number of equilibrium stages.

y value	x value	No. Eq. Stages
0.85	0.85	0
0.85	0.840258969	1
0.840258969	0.828444384	2
0.490471168	0.151519172	9
0.151519172	0.016411729	10
0.016411729	0.002073297	11
0.002073297	0.000286777	12
1.55536E-08	2.18217E-09	18
2.18217E-09	3.06159E-10	19
3.06159E-10	4.29541E-11	20

Table 8. Trial-and-error minimum reflux intersect calculator.

Step 1. Via trial-and-error find x that gives q -line slope as given in user input field.

x (VLE and q -line intercept)	m_{q-line}
0.500000001	152,378,129.76

Slope of line from $x_{Distillate}$ to intercept of the q -line and the VLE line

m_{Rmin}	0.564633929
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Optimum operating conditions

Once the minimum reflux conditions are set, m_{Rmin} is multiplied by the predefined optimum reflux multiplier. The optimum reflux ratio, R , is now defined by the slope of the upper or rectifying operating line. To fully define the system, the point of intersection of the operating line and the q -line must be known, after which the slope of the lower (stripping) operating line can be calculated easily.

Using the operating line slope, m_R , the following equations are employed to find the intersection of the operating line with the q -line at (x_i, y_i) . Based on the upper or rectifying line the equation is:

$$y_i = m_R(x_i - x_D)/x_F \quad (16)$$

This equation has two unknowns, thus, an expression for x_i is formulated from the stripping line:

$$x_i = [(x_F + m_R)/m_{q-line} - x_F]/(m_R/x_D m_{q-line} - 1) \quad (17)$$

When the intersection is found, then the slopes of all the operating lines are calculated by simple algebra, as presented in Table 9. Using the lower and upper operating lines, the previously mentioned algorithm for the McCabe-Thiele diagram is employed to calculate each stage's equilibrium values (Table 10, which shows sample values of the top, middle and bottom groups of stages).

Use of conditional statement

Throughout all of the calculations, incorporating the upper and lower segmentation lines is accomplished by the using the "if" conditional statement. This statement is crucial in calculating the VLE data for the constant y segment of the calculation. Further, an additional "if" statement is necessary at the transition of the stripping and rectifying lines when calculating beyond the values at the intersections of the lines. When the x value of intersection of the operating lines is passed, the spreadsheet uses the stripping operating line to calculate the y values. This is why the segmentation sheet of Table 3 must determine the segmentation transition value of x , as well as that of y .

To find the number of equilibrium stages, note the stage at which the required bottoms conditions are met. For this case, assume that a bottoms x_B value of approximately 0.02 meets the process requirements. The value for Stage 14 is about 0.0135, while that for Stage 15 is about 0.0019, so there are 14 equilibrium stages.

Further calculations are made to find the mean plate efficiency, number of actual stages, the column height, the energy transferred in the reboiler and condenser, and the column diameter (based on the calculated vapor flowrates). These calculations use methods found in standard references, are relatively straightforward, and will only be summarized here.

Table 9. Stripping and rectifying operating lines calculations.

Stripping line equation $y = m_S x + b_S$		
m_S	1.107134375	Must be greater than 1
b_S	0	
Rectifying line equation $y = m_R x + b_R$		
m_R	0.846950893	Cannot be greater than 1
b_R	0.130091741	
Internal reflux rate, $R_{Top} = m_R / (1 - m_R)$ from $m_R = R / (R + 1)$		
R_{Top}	5.533850618	
Point of intersection of operating lines at $R_{Optimum}$		
x		y
0.50000		0.553567187

Table 10. Equilibrium stage calculator.

Number of Equilibrium Stages at R Operating Conditions		
y	x	N_{Eq} No.
0.85	0.85	0
0.85	0.840258969	1
0.616480771	0.411609892	12
0.45570746	0.117412479	13
0.129991391	0.013533436	14
0.014983332	0.001908818	15
0.002113318	0.000292233	16
0.000323541	4.52915E-05	17
3.35118E-21	4.70171E-22	38
5.20543E-22	7.30321E-23	39
8.08564E-23	1.13442E-23	40

Mean plate efficiency and number of stages

The actual number of stages is the number of equilibrium stages divided by the stage efficiency. To determine the efficiency, two empirical methods are used in the program, one described by Van Winkle ((1), p. 557) and

the other by Kister (5). These methods correlate actual efficiencies in petroleum refineries and chemical plants with system parameters such as viscosity.

Van Winkle ((1), p. 557) correlates the stage efficiency, E_o , as a function of the slope of the equilibrium line at a set stage, m , and the viscosity of the mixture, μ , (N-s/m²):

$$E_o = 0.17 - 0.616 \log(m \mu) \quad (18)$$

For binary equilibrium, the average viscosities are calculated using the molar equivalents throughout the height of the column (as the mole fractions change in the liquid segment). For instance, for ethanol/water:

$$\mu_l = \mu_{\text{ethanol}} x_{\text{ethanol}} + \mu_{\text{water}} x_{\text{water}} \quad (19)$$

The slope of the equilibrium line is found by taking the derivative of the equation of the line at a particular point.

All the individual stage efficiencies in the column are calculated, summed and the average efficiency of the entire column is calculated. Then, the theoretical number of stages is divided by the average efficiency to estimate the actual number of stages.

Kister ((5), p. 432) correlates the efficiency with the liquid viscosity and the relative volatility, α :

$$E_o = 0.492 (\alpha \mu_l)^{-0.245} \quad (20)$$

where α is simply:

$$\alpha = y_1(1 - x_1)/x_1(1 - y_1) \quad (21)$$

Be careful when using either of these methods to estimate the stage efficiency. Each may generate a poor prediction of the overall efficiency. A better choice is using the known efficiency for a particular packing or tray (if it is available).

Literature Cited

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4. Treybal, R. E., "Mass Transfer Operations," McGraw Hill, New York (1987).
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Column height and flowrates

The *height* is simply the number of actual stages multiplied by the tray spacing or packing equivalent height. A value of 18 in. is in the spreadsheet.

The *flowrates* are calculated via mass balances. The equations are taken from Treybal ((4), pp. 402-420). Spreadsheets are easily created to perform all of these calculations and convert the results into different systems of units, when desirable.

Reboiler and condenser duties

Calculations of the energy rates of the reboiler and the condenser are made using latent heats and specific heats (when needed). The calculations are as thus performed piecewise, using each chemical species and its respective latent and specific heats.

For example, the bottoms is considered to be made up of 100% water, so its heat duty would be the flowrate of the water vapor rising up from this stage (at 100°C), multiplied by the latent heat of water to create the vapor.

Column diameter

A method described by Kister ((5), pp. 276-279) determines the upper and lower column diameters. Calculations are based on the flooding velocity, the properties of the liquids and vapors, and the fractional hole area in the trays, among other factors. The calculation is straightforward and is not included here.

CEP

JAKE JEVRIC is a chemical engineer at PPG Chemfil (Toronto, Canada; Phone: (416) 247-8897; E-mail: JEVRIC@rogers.com). He is a technical customer service and sales engineer for the automotive-paint applications market. His other professional experience includes jobs in the food and lubricants industries. He has a strong interest in chemical processes design and application, and has developed several spreadsheet applications to simulate chemical design operations. Jevric holds a BE from Ryerson Univ.

MUHAMMAD E. FAYED, P.E. is a professor of chemical engineering at Ryerson Univ. (Dept. of Chemical Engineering, 350 Victoria Street, Toronto, Canada, M5B 2K3; Phone: (416) 979-5217; Fax: (416) 979-5044; E-mail: mfayed@ryerson.ca). Fayed has held parallel professorial appointments in other universities in Canada, the U.S. and the U.K. He heads the Particulate Science and Technology Research Group at Ryerson. He has over 30 years of professional industrial and consulting experience with a number of companies in the U.S., Canada and abroad. For over the last 40 years, he has contributed to the emerging field of crystallization of particulates. Fayed leads numerous professional development courses, seminars and workshops for AIChE, the INTERPHEX Show, and Reed Group companies. He is a member of the advisory editorial board of *Powder & Bulk Engineering*. He has served as a member of the national program committee of AIChE. Fayed holds a PhD from the Univ. of Waterloo, and MSc and BSc degrees from Cairo Univ. He is a Fellow of AIChE, and a Fellow of the Canadian Society of Chemical Engineers (CSCHE).