

Review Of Binary Distillation – McCabe-Thiele By Computer (Part 2)

Two-Feed Distillation

Charles R. Koppany, Retired Chemical Engineer (2009)

In our first paper we presented a BASIC computer program for performing the McCabe-Thiele (1) theoretical stage step-off procedure for a single feed – two product binary distillation column. We emphasized application to close boiling separations such as for splitter fractionators where the operating and equilibrium lines (curves) are generally quite close together and where nearly constant molal overflow prevails.

In this second installment we extend the computer program to handle a **two feed – two product distillation column** and also incorporate the capability to calculate theoretical tray temperatures. In principle this procedure could be applied to any number of feeds where a large number of intermediate operating lines are involved.

Working relationships. First of all, the working relationships used in developing the program will be presented. Our focus here will be restricted to two feed – two (liquid) product distillations. Figure 1 describes the column configuration and nomenclature employed. Here we have three sections – rectifying, middle and stripping.

Equilibrium curve. Once again the binary vapor-liquid coexistence curve is represented by Eqs. 1 and 2.

$$\alpha = Ax^2 + Bx + C \quad (1)$$

$$y = \frac{\alpha x}{1 + (\alpha - 1)x} \quad (2)$$

where x and y are the mole fractions of the light or lower boiling component in the liquid and vapor respectively. α is the relative volatility of the light component with respect to the heavier component (K_{lt}/K_{hvy}). The constants A , B and C of Eq. 1 are fit to a set of isobaric binary VLE data, which hopefully cover the full range of composition for the binary mixture to be distilled. If VLE data are not available, then α values for the fit will have to be generated by some other means such as from a generalized correlation which is believed to give reasonably accurate predictions. Tray-by-tray temperatures can also be calculated if we also have the bubble point (T - x) coexistence curve as well. This curve can be well represented by a similar equation form as used for the relative volatility, namely,

$$T = E x^2 + F x + G \quad (3)$$

Since Eqs. 1 and 3 are parabolic in nature, we must be sure that neither a maximum or minimum (extrema) occurs in either one over the range $0 \leq x \leq 1$.

The q-lines. For a two-feed distillation column there will be two q-lines:

$$\text{Top Feed} \quad y = \frac{q_1}{q_1 - 1} x + \frac{x_{F1}}{1 - q_1} \quad (4)$$

$$\text{Btm. Feed} \quad y = \frac{q_2}{q_2 - 1} x + \frac{x_{F2}}{1 - q_2} \quad (5)$$

In either equation q is defined as the moles of liquid introduced to the section immediately below the feed tray in question resulting from each mole of feed. It basically characterizes the phase condition (mole fraction liquid) of the feed as it enters the column ie.

$$\text{Top feed} \quad q_1 = \frac{L' - L}{F_1} \quad (6)$$

$$\text{Btm feed} \quad q_2 = \frac{\bar{L} - L'}{F_2} \quad (7)$$

Below is listed a summary of the numerical limits for q depending upon the various phase conditions of the entering feed:

<u>Phase Condition</u>	q
Cold feed	> 1
Bubble pt. feed	$= 1$
Two phase feed	$0 < q < 1$
Dew pt. feed	$= 0$
Superheated vapor feed	< 0

Minimum reflux ratio. The minimum reflux ratio $(L/D)_{\min}$ is determined in the same fashion as for a single feed column. It is calculated by finding the point where the top q -line intersects the equilibrium curve. This is Point a shown on Figure 2. Eqs. 1,2 and 4 are combined to eliminate y , and the resulting expression is solved by trial and error substitution to give x' , the liquid mole fraction at Point a. y' is then calculated from Eq. 2 or 4. The minimum reflux ratio is then given by,

$$R_M = \left(\frac{L}{D} \right)_M = \frac{x_D - y'}{y' - x'} \quad (8)$$

Upper operating line. The equation of the upper operating line (UOL) for constant molal overflow is

$$y = \frac{R}{R+1}x + \frac{x_D}{R+1} \quad (9)$$

It is represented as line segment b-c on Figure 2. Here R is the actual operating reflux ratio (L/D) and is established by some specified percentage over the minimum, R_M . In order to determine the intersection point of the UOL and the upper q-line (Point b), it is necessary to equate Eqs. 4 and 9. The resulting expression is solved directly for x'' , the liquid composition at this point of intersection. y'' is then determined by either Eq. 4 or 9.

Middle operating line. Before the middle operating line (MOL) can be established, an overall material balance must be performed in order to establish the overhead and bottoms product rates. The result is,

$$D = \frac{x_{F1}F_1 + x_{F2}F_2 - x_B(F_1 + F_2)}{x_D - x_B} \quad (10)$$

$$B1 = F_1 + F_2 - D \quad (11)$$

Here both the upper and lower feed rates and their compositions (x_{F1} , x_{F2}) must be specified.

Next the liquid and vapor flow rates in the rectifier (upper section) are determined by,

$$L = (R)(D) \quad (12)$$

$$V = L + D \quad (13)$$

The liquid and vapor flow rates in the middle section are established by using the definition of q_1 for the top feed (Eq. 6) combined with a simple material balance around the top feed tray itself. The result is,

$$L' = L + q_1 F_1 \quad (14)$$

$$V' = V - (1 - q_1) F_1 \quad (15)$$

Now the slope and intercept for the MOL can be calculated from Eqs. 16-18.

$$y = (M_2)x + I_2 \quad (16)$$

where
$$M_2 = \frac{L'}{V'} \quad (17)$$

and
$$I_2 = y'' - (M_2)x'' \quad (18)$$

Eq. 18 is valid because the MOL is connected at one end by x'', y'' which is point b on Figure 2.

The intersection of the bottom q-line and the MOL is established by equating Eqs. 5 and 16. The solution of these equations yields Point d (x''', y''') on Figure 2.

Lower operating line. The lower operating line (LOL) is now very easily determined.

It is simply the line segment d-e with slope given by,

$$\text{slope} = M_3 = \frac{y''' - x_B}{x''' - x_B} \quad (19)$$

$$\text{Intercept} = I_3 = x_B(1 - M_3) \quad (20)$$

Finally, the vapor-liquid flow traffic in the stripping (bottom) section can be conveniently computed via Eqs. 21 and 22,

$$\bar{L} = L' + q_2 F_2 \quad (21)$$

$$\bar{V} = V' - (1 - q_2) F_2 \quad (22)$$

The program. The computer program, employing all of the above relationships for performing the McCabe-Thiele graphical construction procedure numerically, is listed in Table 1. The program is written in QBASIC and has been adapted to run on a PC computer (MS-DOS).

Lines 40-90 request inputs for the coefficients A, B and C of Eq. 1 and E, F and G for Eq. 3. These equations along with Eq. 2 establish the equilibrium y-x curve and T-x (isobaric) bubble point curve. Line 97 provides a command to read in values for the distillate, bottoms, upper and lower feed compositions, specification of q_1 and q_2 for upper and lower feeds, and the upper and lower total feed flow rates (mols/hr). All compositions are inputted as mole fraction of the light or lower-boiling component. In Lines 220-230 the values of q_1 and q_2 are set to 0.999999 if either of them were originally inputted as 1.0 (saturated liquid feed). This specification avoids the occurrence of a singularity in either Eq. 4 or 5. Line 240 requires an inputted value for the ratio of operating to minimum reflux ratio (percentage over minimum), and lines 300 and 310 represent the componential and overall material balances for computing the distillate and bottoms flow rates (mols/hr).

Lines 340-650 employ all of the equations presented in the text to establish the minimum reflux ratio, the slopes and intercepts of the upper (rectifying), middle and lower (stripping) operating lines, and the mutual points of intersection of the operating lines and the q-lines.

The remainder of the program is devoted to stage-by-stage computations based on the previously established equilibrium curve and operating lines. Starting with x_B , the specified bottoms liquid composition exiting the reboiler, the equilibrium vapor composition y_B exiting the reboiler and the reboiler temperature are both calculated. Obviously, this point (x_B, y_B) lies on the equilibrium curve as shown on Figure 3. From the value of y_B the composition of the liquid exiting the bottom theoretical tray of the column itself is calculated from the equation of the LOL (Eqs. 19, 20). Graphically these calculated points form the bottom triangle on Figure 3. This is the first theoretical stage in the calculation sequence and represents the reboiler. In this way the program works its way up the column following the LOL and equilibrium curve until the equilibrium vapor composition first exceeds y''' (Point d). At this point the program is ready to switch over to the MOL. The number of theoretical stages (triangles) recorded at this point correspond to the exact number for the stripping section, including the lower (bottom) feed tray and reboiler. As a result, the bottom feed tray is easily located. It is basically represented by the triangle with one corner on the LOL and the other on the MOL. Next the program proceeds to perform stage-by-stage computations up the "middle section" by following the MOL. When the equilibrium vapor composition just exceeds y'' (Point b), the program then switches over to the UOL (rectifier). This latter step-off (triangle) represents the upper (top) feed tray. Finally, the program proceeds up the UOL and proceeds until the vapor composition becomes equal to or greater than x_D , the specified overhead product composition. At this point the program stops and notes the

last stage computed as the top rectifying tray. If it is required that the distillate exist as a dewpoint vapor, this top theoretical tray represents a partial condenser.

Example. Let us consider a fractionator which must be roughly sized to distill two separate feed streams into a single bubble point liquid distillate and a single bubble point liquid bottoms product. The fractionation is to be conducted at a mean operating pressure of 150 psia and involves a separation of the binary system propylene-1-butene. The feed compositions and desired product purities (specs.) are as follows:

Stream Mole fractions

	<u>Top Feed</u>	<u>Btm. Feed</u>	<u>Distillate</u>	<u>Bottoms</u>
Propylene	0.60	0.30	0.95	0.05
1-butene	<u>0.40</u>	<u>0.70</u>	<u>0.05</u>	<u>0.95</u>
	1.00	1.00	1.00	1.00

Feed Conditions:

	<u>Mols/hr Total Flow</u>	<u>q</u>
Top Feed	100	1.0 BP liquid
Btm. Feed	100	0.0 DP Vapor

It is desired to operate this column at a reflux ratio that is 100 % over the minimum ($R/R_M = 2.0$). We are asked to calculate (estimate) the following design parameters:

1. Tray-by-tray vapor/liquid compositions and temperature profiles
2. Location of both feed trays
3. Total number of theoretical trays required

We can safely assume here that such a nearly ideal and close-boiling system can be fractionated under nearly constant molal overflow conditions. As a result, no heat balancing is required in the design calculations, and the McCabe-Thiele method would be very appropriate to use.

At low to moderate pressures the VLE for any mixture of hydrocarbons from the same homologous series, such as the olefinic hydrocarbons, closely obeys Raoult's and Dalton's laws. For these systems, the vapor-liquid equilibrium ratio is simply equal to the ratio of the component vapor pressure to the total system pressure.

$$K_i = \frac{y_i}{x_i} = \frac{P_i^{VP}}{P} \quad (23)$$

where x_i is the mole fraction of species i in the liquid, and y_i is the mole fraction of i in the vapor phase. The isobaric T-x-y coexistence curves for a binary system obeying Raoult's and Dalton's laws are simply computed from the relations,

$$x_1 = \frac{1 - K_2}{K_1 - K_2} \quad (24)$$

$$y_1 = K_1 x_1 \quad (25)$$

where 1 refers to the lower boiling component and 2 to the higher boiler. The values of K_1 and K_2 are simply computed from Eq. 23.

For the system propylene-1-butene at 150 psia, the pure component saturation temperatures of propylene and 1-butene are 69 deg. F and 163 deg. F respectively. At intermediate temperatures, the isobaric T-x-y diagram is readily computed from Eqs. 23-25. Vapor pressures were read from charts present in the Data Book of the American

Petroleum Institute (2). The tables below summarize the results of these calculations:

<u>T, °F</u>	<u>P, Psia</u>	<u>VP_i Psia</u>		<u>K_i = VP_i/P</u>		<u>m.f. C₃H₆</u>	
		<u>C₃H₆</u>	<u>C₄H₈</u>	<u>C₃H₆</u>	<u>C₄H₈</u>	<u>x</u>	<u>y</u>
69	150	150	-	1.0	-	1.0	1.0
80	150	176	47	1.173	0.313	0.798	0.937
100	150	227	63	1.513	0.420	0.530	0.803
120	150	283	84	1.887	0.560	0.332	0.626
140	150	353	110	2.353	0.733	0.165	0.387
150	150	396	125	2.640	0.833	0.092	0.244
163	150	-	150	-	1.0	0.0	0.0

Relative volatilities, $\alpha = K_{C_3}/K_{1-C_4}$:

<u>T, °F</u>	<u>x-C₃H₆</u>	<u>Equilibrium Ratio, K_i</u>		
		<u>C₃H₆</u>	<u>C₄H₈</u>	<u>α</u>
69	1.0	1.0	-	-
80	0.798	1.173	0.313	3.745
100	0.530	1.513	0.420	3.603
120	0.332	1.887	0.560	3.369
140	0.165	2.353	0.733	3.209
150	0.092	2.640	0.833	3.168
163	0.0	-	1.0	-

The values of α and T above were fitted to Eqs. 1 and 3 respectively using the linear regression routine available in the spreadsheet program EXCEL. The following results were obtained:

$$\alpha = -0.3956x^2 + 1.212849x + 3.037908 \quad (26)$$

$$T = 52.7799x^2 - 146.474x + 162.9095 \quad (27)$$

Here x is the mole fraction propylene in the liquid, and T is the operating temperature in degrees F. The fits are excellent. Eq. 26 produces an absolute average deviation of 0.65 percent for the prediction of α , and Eq. 27 yields an absolute average deviation of only 0.2 percent for the prediction of temperature.

The detailed output and tray-by-tray calculations generated from the QBASIC program are listed in Table 2. The first portion of the output summarizes all of the input data. The coefficients of Eqs 1 and 3 are printed followed by the product and feed compositions, values of q_1 and q_2 , R/R_{\min} , top and bottom feed flow rates, and the calculated distillate and bottoms flow rates based on Eqs. 10 and 11. Next the minimum and operating reflux ratios are printed along with the intersection points (x,y compositions) between the q and operating lines.

The remaining output lists stage-by-stage liquid and vapor mole fractions of propylene in 1-butene, relative volatilities and temperatures ($^{\circ}$ F). There are four stripping trays with tray 1 being the reboiler and tray no. 4 being the bottom or lower feed tray. Trays 5 through 9 are present in the middle section where tray 9 is the top or upper feed tray. The

three remaining trays (10 to 12) comprise the rectifier or upper section of the column.

The vapor composition exiting tray 12 is 0.984 mole fraction propylene which exceeds the value for the distillate composition spec. (0.95 m.f. propylene). If a vapor distillate is required, then tray 12 represents a partial condenser.

In summary we have the following theoretical stage profile numbered from the bottom up:

<u>Tray</u>	
12	Partial condenser or top tray
11	
10	
9	Top feed tray
8	
7	
6	
5	
4	btm. Feed tray
3	
2	
1	reboiler

Finally, the vapor-liquid flow rate traffic in all three column sections are printed.

If the column employs a total condenser (BP liquid distillate), there will be:

1 reboiler (theoretical stage)

11 theoretical trays in the column

If the column requires a dew point vapor distillate, the stage breakdown for this particular column design would:

1 reboiler and 1 (partial) condenser (2 theoretical stages)

10 theoretical trays in the column

Figure 4 shows the actual McCabe-Thiele graphical construction for this particular example. The computer results agree very nicely with the graphical solution.

LITERATURE CITED

1. McCabe, W.L. and E.W. Thiele, Ind. Eng. Chem., 17, No. 6, p. 605 (1925).
2. Technical Data Book of the American Petroleum Institute (API)

NOMENCLATURE

A, B and C	Coefficients to Eq. 1
B ₁	Bottoms flow rate, mols/hr
D	Distillate flow rate mols/hr
E, F and G	Coefficients to Eq. 3
F ₁	Top feed flow rate, mols/hr
F ₂	Bottom feed flow rate, mols/hr
I ₁	Intercept of the rectifying (upper) operating line (UOL)
I ₂	Intercept of the middle operating line (MOL)
I ₃	Intercept of the lower operating line (LOL or stripper)
L	Liquid reflux or rectifier liquid flow rate, mols/hr
L'	Liquid flow rate in the middle section, mols/hr
L(bar)	Liquid flow rate in the stripping (lower) section, mols/hr
M ₁	Slope of the rectifying or upper operating line (UOL)
M ₂	Slope of the middle operating line (MOL)
M ₃	Slope of the lower operating line (LOL or stripper)

NOMENCLATURE Cont.-

N_{theo} or N	Total number of theoretical trays at operating reflux ratio, R
q_1	Moles of liquid introduced to middle section of column resulting from introduction of each mole of top feed
q_2	Moles of liquid introduced to stripping (lower) section of column resulting from introduction of each mole of bottom feed
R	Operating reflux ratio at some percentage over the minimum, L/D
R_M	Minimum reflux ratio (infinite stages) or $(L/D)_{\text{min}}$
x	Mole fraction of lower boiling or light component in the liquid
x'	Liquid mole fraction of lower boiling or light component at the intersection of the top q -line and the equilibrium curve
x''	Liquid mole fraction of the lower boiling or light component at the intersection of the top q -line and the upper operating line
x'''	liquid mole fraction of the lower boiling or light component at the intersection of the bottom q -line and the middle operating line
x_B	Mole fraction of the lower boiling or light component in the bottoms product
x_D	Mole fraction of the lower boiling or light component in the distillate product
x_{F1}	Mole fraction of the lower or light component in the top feed
x_{F2}	Mole fraction of the lower or light component in the bottom feed

NOMENCLATURE Cont.-

y	Mole fraction of the lower boiling or light component in the vapor
y'	Vapor mole fraction of the lower boiling or light component at the intersection of the top q-line and the equilibrium curve
y''	Vapor mole fraction of the lower boiling or light component at the intersection of the top q-line and the upper operating line
y'''	Vapor mole fraction of the lower boiling or light component at the intersection of the bottom q-line and the middle operating line
y_B	Mole fraction of the lower boiling or light component in the vapor exiting the reboiler and entering the bottom theoretical tray
α	Relative volatility of the lower boiling component with respect to the higher boiling component or ratio of their equilibrium constants, K_{lt}/K_{hvy}

Table 1

Listing of the QBASIC Program MCTH2F1.BAS

```

10 LPRINT "2-FEED MCCABE/THIELE BINARY DISTILLATION"
20 LPRINT
40 READ A, B, C
60 LPRINT "A", "B", "C"
70 LPRINT A, B, C
80 LPRINT
85 READ E, F, G
88 LPRINT "E", "F", "G"
90 LPRINT E, F, G
95 LPRINT
97 READ XD, XB, Z1, Z2, Q1, Q2, F1, F2
100 LPRINT
110 LPRINT "DISTILLATE COMPOSITION, XD="; XD
130 LPRINT "BOTTOMS COMPOSITION, XB="; XB
150 LPRINT "TOP FEED COMPOSITION, Z1="; Z1
170 LPRINT "BTM FEED COMPOSITION, Z2="; Z2
190 LPRINT "TOP FEED THERMAL CONDITION, Q1="; Q1
210 LPRINT "BTM FEED THERMAL CONDITION, Q2="; Q2
220 IF Q1 = 1 THEN Q1 = .999999
230 IF Q2 = 1 THEN Q2 = .999999
240 INPUT "R/RMIN="; R0
250 LPRINT "R/RMIN="; R0
270 LPRINT "TOP FEED RATE, M/H="; F1
290 LPRINT "BTM FEED RATE, M/H="; F2
300 D = (Z1 * F1 + Z2 * F2 - XB * (F1 + F2)) / (XD - XB)
310 B1 = F1 + F2 - D
320 LPRINT "DISTILLATE RATE, M/H="; D
330 LPRINT "BTMS RATE, M/H="; B1
340 X0 = .5
350 A0 = A * X0 * X0 + B * X0 + C
360 Y0 = (A0 * X0) / (1 + (A0 - 1) * X0)
370 X1 = (Y0 - Z1 / (1 - Q1)) * (Q1 - 1) / Q1
380 E1 = ABS(X1 - X0)
390 IF E1 < .0001 THEN 420
400 X0 = X1
410 GOTO 350
420 RM = (XD - Y0) / (Y0 - X0)
430 LPRINT
440 LPRINT "RMIN="; RM
450 R = R0 * RM
460 LPRINT
470 LPRINT "RACTUAL="; R
480 M1 = R / (R + 1)
490 I1 = XD / (R + 1)
500 X2 = (Z1 / (1 - Q1) - I1) / (M1 - Q1 / (Q1 - 1))
510 Y2 = M1 * X2 + I1
520 LPRINT

```

```

530 LPRINT "INTERSECTION OF UPPER O.L. AND TOP FEED Q-LINE, X,Y=";
X2, Y2
540 L = R * D
550 V = L + D
560 LPR = L + Q1 * F1
570 VPR = V - (1 - Q1) * F1
580 M2 = LPR / VPR
590 I2 = Y2 - M2 * X2
600 X3 = (Z2 / (1 - Q2) - I2) / (M2 - Q2 / (Q2 - 1))
610 Y3 = M2 * X3 + I2
620 LPRINT
630 LPRINT "INTERSECTION OF MIDDLE O.L. AND BTM FEED Q-LINE,X,Y=";
X3, Y3
640 M3 = (Y3 - XB) / (X3 - XB)
650 I3 = XB * (1 - M3)
660 N = 1
662 TB = E * XB * XB + F * XB + G
670 AL = A * XB * XB + B * XB + C
680 YB = (AL * XB) / (1 + (AL - 1) * XB)
690 LPRINT
700 LPRINT "STAGE", "X.LIQUID", "Y,VAPOR", "ALPHA", "TEMP,F"
710 LPRINT N, XB, YB, AL, TB
720 Y = YB
730 X = (Y - I3) / M3
732 T = E * X * X + F * X + G
740 AL = A * X * X + B * X + C
750 Y = (AL * X) / (1 + (AL - 1) * X)
760 N = N + 1
770 LPRINT N, X, Y, AL, T
780 IF Y > Y3 THEN 800
790 GOTO 730
800 LPRINT "TOP OF STRIPPING SECTION"
810 LPRINT
820 LPRINT
830 X = (Y - I2) / M2
832 T = E * X * X + F * X + G
840 AL = A * X * X + B * X + C
850 Y = (AL * X) / (1 + (AL - 1) * X)
860 N = N + 1
870 LPRINT N, X, Y, AL, T
880 IF Y > Y2 THEN 900
890 GOTO 830
900 LPRINT "TOP OF MIDDLE SECTION"
910 LPRINT
920 LPRINT
930 X = (Y - I1) / M1
932 T = E * X * X + F * X + G
935 IF X > XD THEN 990
940 AL = A * X * X + B * X + C
950 Y = (AL * X) / (1 + (AL - 1) * X)
960 N = N + 1
970 LPRINT N, X, Y, AL, T
980 GOTO 930
990 LPRINT "TOP OF RECTIFYING SECTION"

```

```
1000 LPRINT
1005 LPRINT
1010 LBAR = LPR + Q2 * F2
1015 VBAR = VPR - (1 - Q2) * F2
1020 LPRINT "L AND V RATES, M/H, IN RECTIFUER="; L, V
1025 LPRINT
1030 LPRINT "L AND V RATES, M/H, IN MIDDLE SCT="; LPR, VPR
1035 LPRINT
1040 LPRINT
1045 LPRINT "L AND V RATES, M/H, IN STRIPPER="; LBAR, VBAR
2000 DATA -0.3956,1.212849,3.037908
2010 DATA 52.7799,-146.474,162.9095
2020 DATA 0.95,0.05,0.6,0.3,1,0,100,100
5000 END
```

Table 2

Output For Example

2-FEED MCCABE/THIELE BINARY DISTILLATION

A	B	C
-.3956	1.212849	3.037908
E	F	G
52.7799	-146.474	162.9095

DISTILLATE COMPOSITION, $X_D = 0.95$
 BOTTOMS COMPOSITION, $X_B = 0.05$
 TOP FEED COMPOSITION, $X_{F1} = 0.6$
 BTM FEED COMPOSITION, $X_{F2} = 0.3$
 TOP FEED THERMAL CONDITION, $Q_1 = 1$
 BTM FEED THERMAL CONDITION, $Q_2 = 0$
 $R/R_{MIN} = 2$
 TOP FEED RATE, $M/H = 100$
 BTM FEED RATE, $M/H = 100$

DISTILLATE RATE, $M/H = 88.89$
 BTM RATE, $M/H = 111.11$

$R_{MIN} = 0.43094$
 $R_{ACTUAL} = 0.86188$

INTERSECTION OF UPPER O.L AND TOP FEED Q-LINE, $X, Y = 0.60 \quad 0.7880$

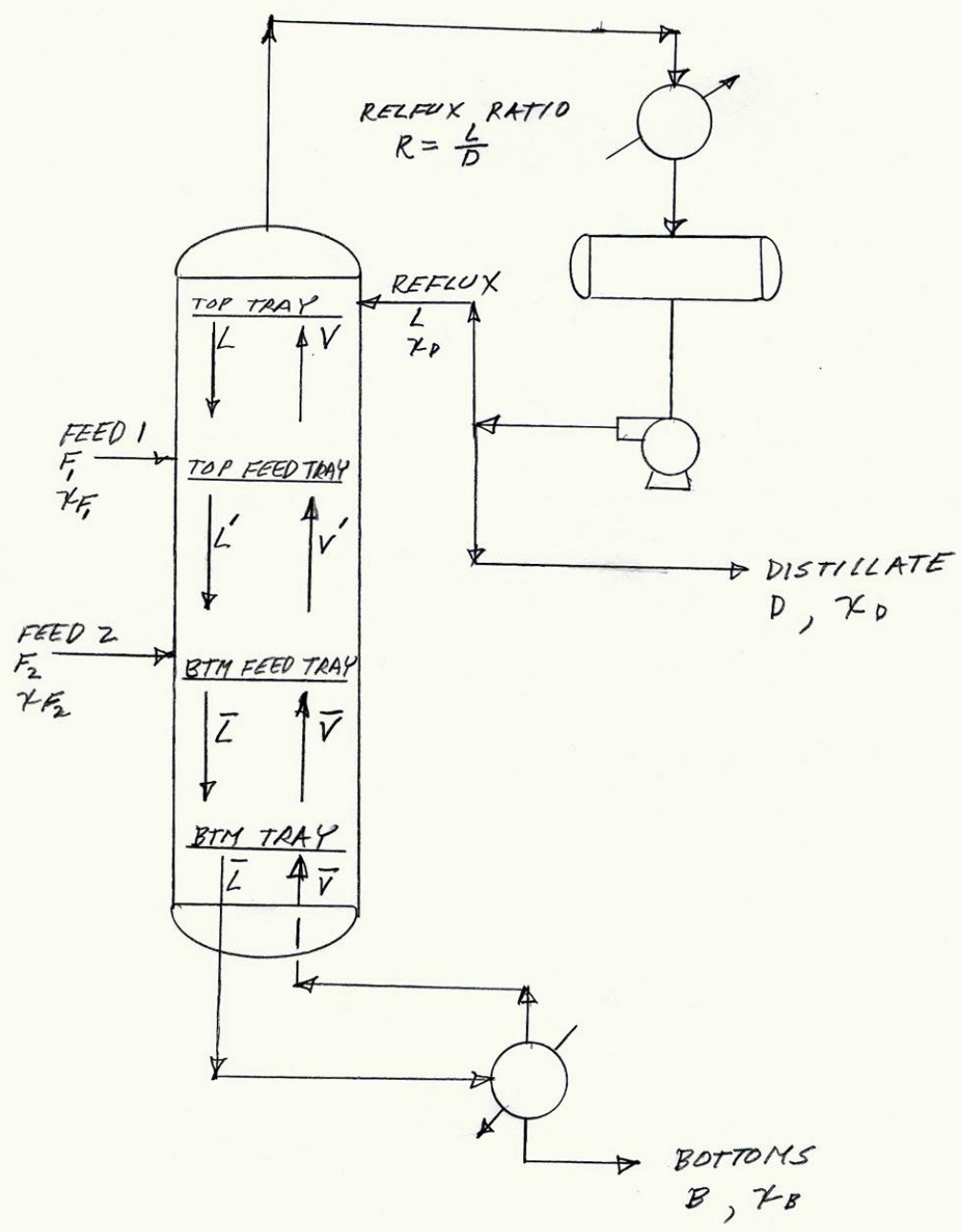
INTERSECTION OF MIDDLE O.L. AND BTM FEED Q-LINE, $X, Y = 0.1427, 0.30$

STAGE	X-LIQUID	Y-VAPOR	ALPHA	Temp., F	
1	0.05	0.14018	3.0976	155.8	REBOILER
2	0.0834	0.2221	3.136	151.1	
3	0.1138	0.2894	3.171	146.9	
4	0.1388	0.3402	3.199	143.6	BTM FEED TRAY
TOP OF STRIPPING SECTION					
5	0.1803	0.4165	3.244	138.2	
6	0.2519	0.5277	3.318	129.4	
7	0.3561	0.6541	3.420	117.4	
8	0.4745	0.7609	3.524	105.3	
9	0.5746	0.8296	3.604	96.2	TOP FEED TRAY
TOP OF MIDDLE SECTION					
10	0.6899	0.8913	3.686	87.0	
11	0.8233	0.9461	3.768	78.1	
12	0.9416	0.9841	3.829	71.8	CONDENSER
TOP OF RECTIFYING SECTION					

Table 2 Continued-

L AND V RATES, M/H, IN RECTIFIER = 76.6	165.5
L AND V RATES, M/H, IN MIDDLE SCT. = 176.6	165.5
L AND V RATES, M/H, IN STRIPPER = 176.6	65.5

FIGURE 1
COLUMN NOMENCLATURE



2 Feed Distillation

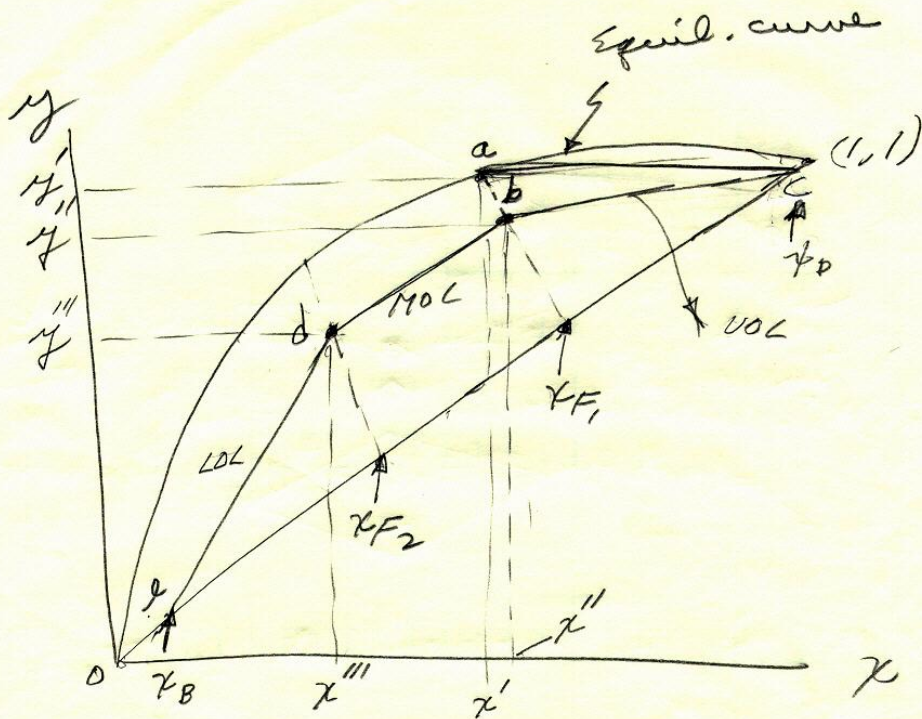


Fig. 2 - Determination of o.c.'s.

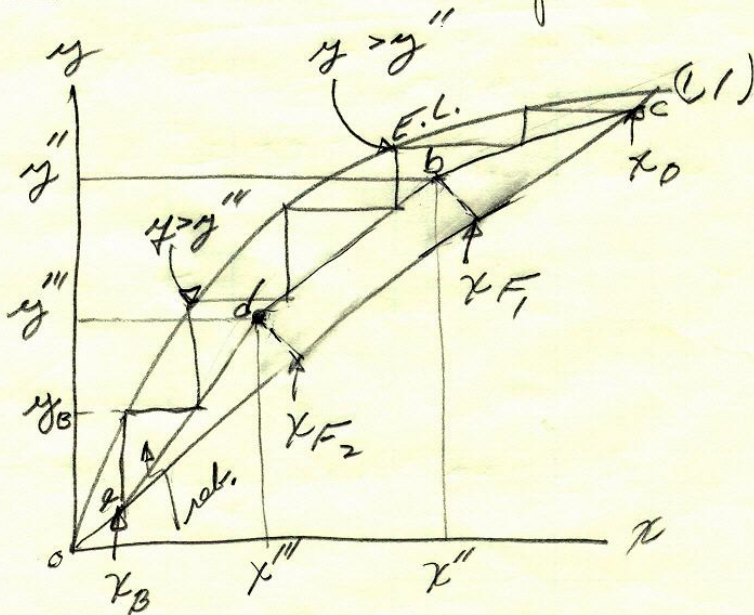


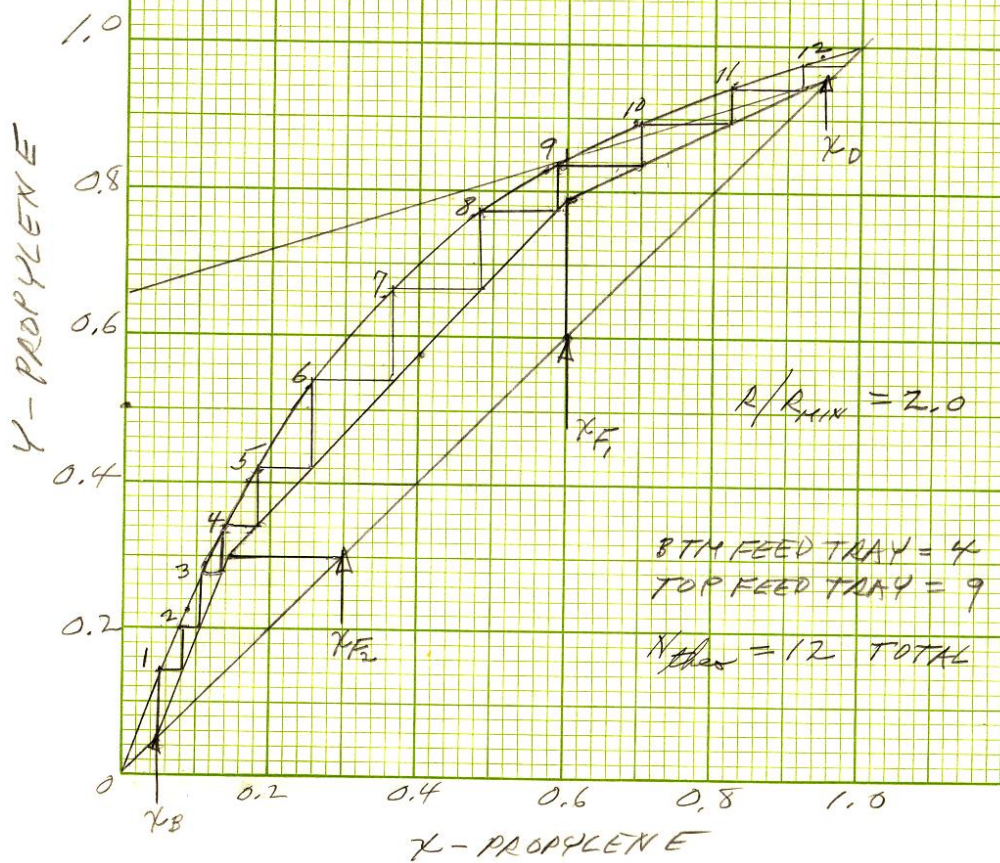
Fig. 3 - Stage - by - Stage stepoff.

CRK 2-7-09

FIGURE 4

GRAPHICAL MCCABE-THIELE
CONSTRUCTION FOR EXAMPLE

SYSTEM: PROPYLENE/1-BUTENE
 $P = 150 \text{ PSIA}$



AUTHOR'S BACKGROUND

Dr. Charles R. Koppany is a retired chemical engineer formerly employed by C F Braun & Co/ Brown & Root, Inc. from 1965 to 1994. While at Braun he served in both the Research and Process Engineering departments. Dr. Koppany has also done part-time teaching in the Chemical Engineering Departments at Cal Poly University Pomona and the University of Southern California. He holds B.S., M.S. and PhD degrees in Chemical Engineering from the University of Southern California and is a registered professional engineer (Chemical) in the state of California.

crkoppany@aol.com

