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Rigorous Computation of Binary Distillation Systems

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ABSTRACT

A computation algorithm for the distillation of binary systems based on the concepts involved in the graphical Ponchon–Savarit method is introduced. Equilibrium and enthalpy data are estimated by the UNIQUAC activity coefficient equation. All the parameters of the system (temperatures, enthalpies, flow rates, and composition of the streams in each stage) are computed with much higher accuracies than those obtained by the graphical method.

INTRODUCTION

The Ponchon–Savarit graphical method (1, 2) has been extensively used for calculating the number of theoretical stages in a distillation tower used for separating a binary liquid mixture. This method has widespread acceptance over the McCabe–Thiele method since no simplifying assumptions have to be introduced, especially the need for the constant phase flow ratio assumption. The main advantage of this rigorous method is its ability to estimate the minimum reflux ratio, the condenser duty, and the reboiler load. In addition, the number of stages, feed stage position, enthalpies, and composition of the streams from stage to stage can also be evaluated. This method makes use of enthalpy–concentration and equilibrium diagrams for the binary vapor–liquid mixture under consideration.

The enthalpy–concentration diagram takes into account latent heats, heats of solutions or mixing, and sensible heats of the components of the mixture. The difficulty with this method is that heats of solution are not usually available. Govind (3) developed an analytical form of the Ponchon–Savarit method for systems with straight enthalpy–composition phase lines. The main outcome of this technique is its ability to find the minimum reflux ratio analytically rather than by the trial-and-error graphical method. However, Govind (3) stated that his method gave results identical to those obtained by the McCabe–Thiele method when the enthalpy–composition phase lines were linear. However, the enthalpy data are generally of such a large magnitude that their graphical representation may be liable to compounded errors, which may be reflected in the results obtained.

In the present work the enthalpy–composition and equilibrium data are correlated by simple polynomials. The computation technique presented is based on solving the materials and energy balances on each stage in the tower with the correlations developed. All the parameters which can be estimated by the graphical method can be produced by this algorithm with higher degrees of accuracy.

THE PONCHON–SAVARIT METHOD

The Ponchon–Savirat method is usually applied to binary distillation and ternary extraction systems. It employs a graphical technique based on mass and energy balances which are represented on the enthalpy concentration diagram of the system.

An enthalpy concentration diagram (H–x–y) can be generated by calculating the compositions and enthalpies of the coexisting liquid and vapor phases of the system at equilibrium. The H–x data for the liquid and the H–y data for the vapor are then plotted on a rectangular diagram. The well-known inverse level rule applies to this diagram. The graphical technique will not be discussed here. However, it can be found elsewhere (4).

EQUILIBRIUM AND ENTHALPY RELATIONSHIPS

Vapor–liquid equilibrium data may be estimated from equations of state or from activity coefficient correlations (4).

In the present work, the equilibrium ratio for any component i at low pressures is estimated by

$$K_i = \gamma_{il} P_{vi} / P \qquad (1)$$

where P is the total pressure of the system. P_{vi} is the saturated vapor pressure which can be estimated by the Antoine equation. γ_{il} is the activity coefficient of component i in the liquid phase. The value of this coefficient is estimated here by the UNIQUAC equation.

Once the values of the equilibrium ratios for a certain system are evaluated, the vapor and liquid concentrations y and x can be correlated by a polynomial of a suitable degree. These polynomials take the forms

$$y = P_1(x) \tag{2}$$

$$x = P_2(y) \tag{3}$$

The enthalpy of saturated liquid solutions includes both the sensible heat and the heat of mixing:

$$h = \Sigma x_i C_{Li}(t_b - t_0) + H_E \tag{4}$$

where h is the enthalpy of the saturated liquid mixture in energy per mole, C_{Li} is the heat capacity of component i in the mixture in energy per mole degree, t_b is the bubble point of the mixture corresponding to the liquid concentration at the prevailing pressure, t_0 is a reference temperature, and H_E is the molar excess enthalpy of solution at t_0 and the prevailing concentration referred to the pure liquid components in energy per mole solution. The excess enthalpy of solution can be estimated using the UNIQUAC equation parameters (5):

$$H_E = R\Sigma\{[q'_i x_i)/(\Sigma\theta'_j\tau_{ji})](\Sigma\theta'_j a_{ji})\} \tag{5}$$

where R is the gas constant, q'_i is a surface area parameter of the molecule i, and θ'_i, τ_{ji}, and a_{ji} are parameters of the UNIQUAC equation, defined as follows:

$$\theta'_i = (x_i q'_i)/(\theta x_i q'_i) \tag{6}$$

$$\tau_{ji} = \exp[-a_{ji}/T] \tag{7}$$

a_{ji} is an interaction parameter between the pairs of components j and i. The value of this parameter is obtained by fitting the UNIQUAC model to the experimental data. Values of these binary interaction parameters for quite a number of components are available elsewhere (5).

The enthalpy of the liquid mixture estimated can then be correlated by the following polynomial:

$$h = P_3(x) \tag{8}$$

The enthalpy of saturated vapor mixtures can be estimated by the sensible heat and heat of vaporization of the components of the mixture:

$$H = \Sigma y_i\{C_{Li}(t_d - t_0) + \lambda_i\} \tag{9}$$

where H is the saturated vapor enthalpy in energy per mole, t_d is the dew point of the vapor mixture at the specified concentration and pressure, and λ_i is the latent heat of vaporization at the dew temperature t_d in energy per mole.

Once the values of the vapor enthalpies are calculated, they can then be correlated by

$$H = P_4(y) \tag{10}$$

THE PONCHON–SAVARIT TECHNIQUE

The Ponchon technique makes use of the usual material and energy balances on the distillation column and on the rectifying and stripping sections of this column.

Consider a certain feed mixture which is introduced into a column such as that shown in Fig. 1. For a given feed flow rate F with concentration z_F, temperature t_F, enthalpy h_F, and thermal condition q, the material and energy balance equations can be written as

$$F = D + B \tag{11}$$

$$Fz_F = Dz_D + Bx_B \tag{12}$$

$$Fh_F = DQ_D + BQ_B \tag{13}$$

where D and B are the distillate and bottom flow rates, respectively, z_D and x_B are the distillate and bottom concentrations of the more volatile component in mole fractions, and Q_D and Q_B are the net heats in the condenser and reboiler. These quantities are defined by the following relations:

$$Q_D = h_D + Q_C/D \tag{14}$$

$$Q_B = h_B - Q_R/B \tag{15}$$

where h_D and h_B are the enthalpies of the distillate and bottom streams, respectively, in energy per mole, and Q_C and Q_R are the heat removed from the condenser and the heat supplied to the reboiler, respectively, in energy per unit time. Solving Eqs. (11), (12), and (13) gives

$$D/B = (z_F - x_B)/(z_D - z_F) = (h_F - Q_B)/(Q_D - h_F) \tag{16}$$

Equation (16) represents the overall material and energy balances of the column. It can be represented by a straight line (dfb) on the Ponchon diagram, shown in Fig. 2. This line connects the condenser to the bottom through the feed point.

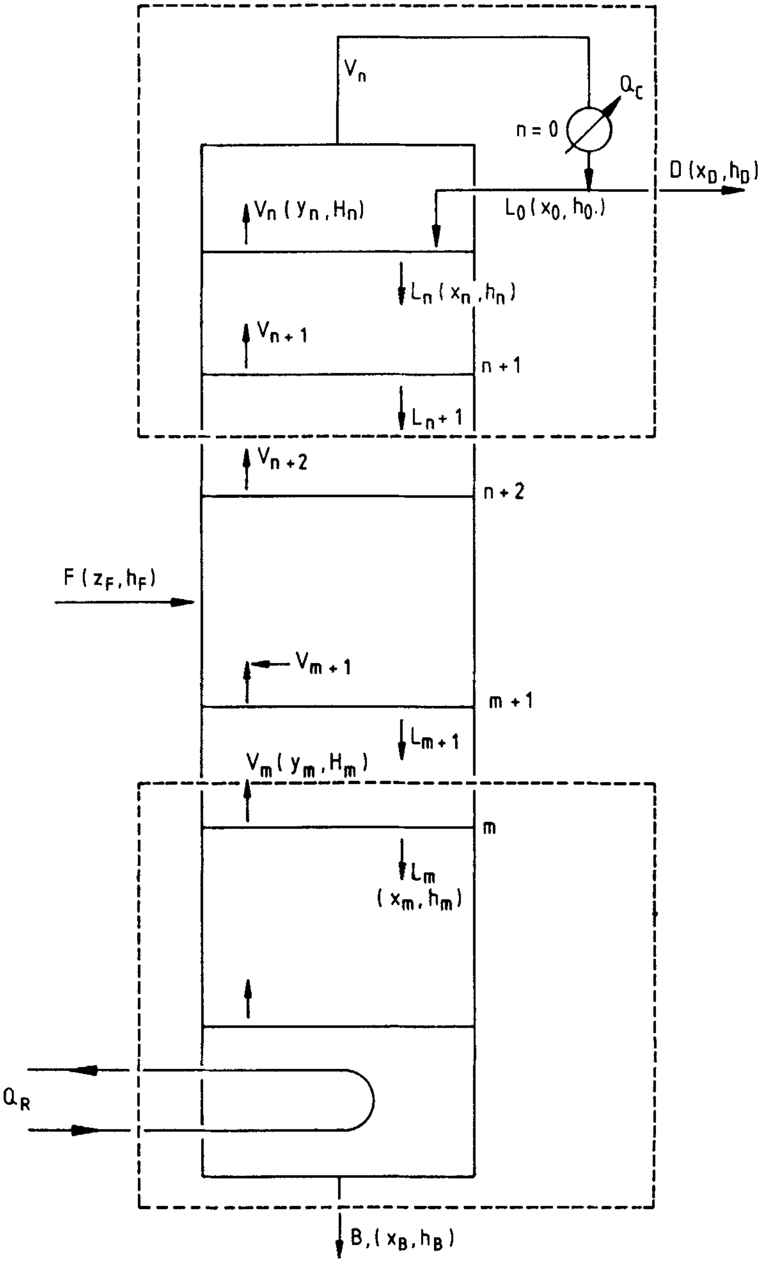

FIG. 1 Distillation column with total condenser.

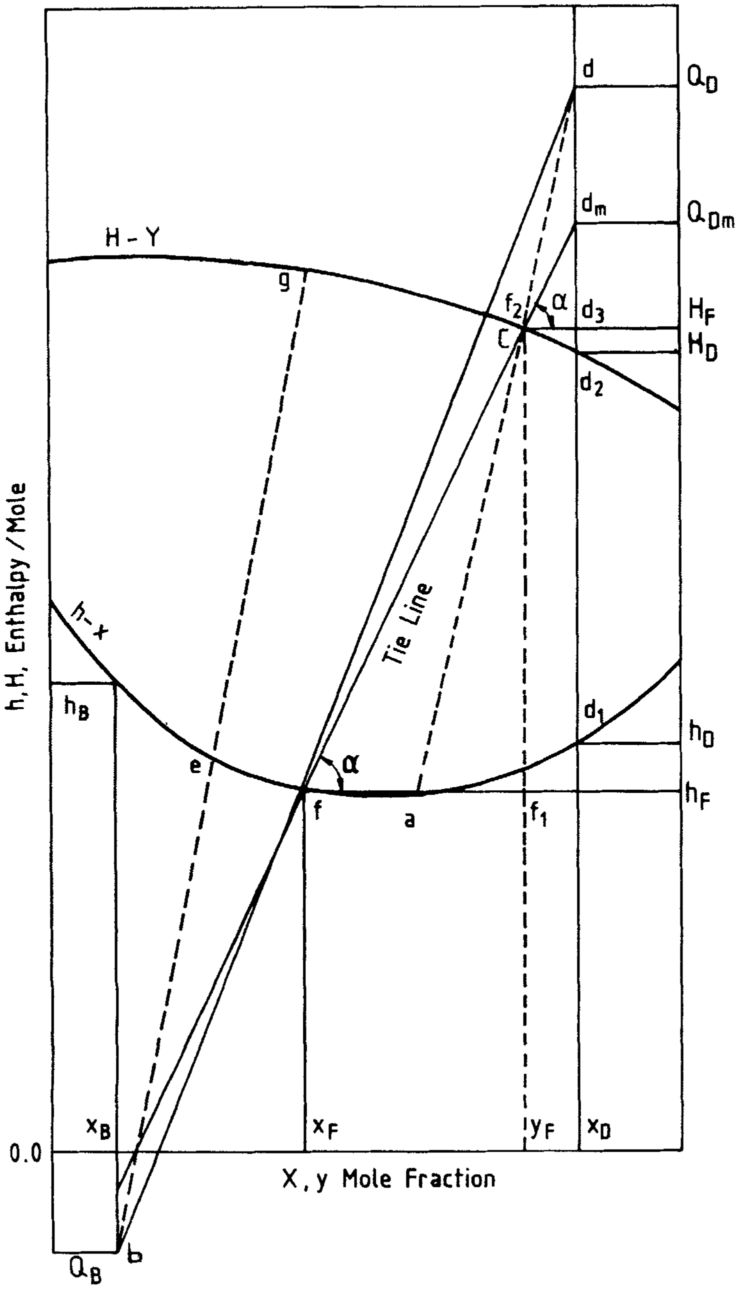

FIG. 2 Ponchon diagram.

Similar balances can be applied on the condenser to get

$$L_n/V_{n+1} = (z_D - y_{n+1})/(z_D - x_n) = (Q_D - H_{n+1})/(Q_D - h_n) \quad (17)$$

Equation (17) can be rearranged in a different way (6) to give

$$L_n/D = (z_D - y_{n+1})/(y_{n+1} - x_n) = (Q_D - H_{n+1})/(H_{n+1} - h_n) \quad (18)$$

Both Eqs. (17) and (18) represent any straight line (dCa) in the rectifying section where Point d represents the condenser (z_D, Q_D), Point c represents the vapor flow rate of Stage $n + 1$, $V_{n+1}(y_{n+1}, H_{n+1})$, and Point a represents the liquid flow rate of Stage n, $L_n(x_n, h_n)$.

Similar equations can be derived from the material and energy balances of the reboiler to represent the stripping section stages.

$$L_m/V_{m+1} = (y_{m+1} - x_B)/(x_m - x_B) = (H_{m+1} - Q_B)/(h_m - Q_B) \quad (19)$$

$$L_m/B = (y_{m+1} - x_B)/(y_{m+1} - x_m) = (H_{m+1} - Q_B)/(H_{m+1} - h_m) \quad (20)$$

These two equations are represented on the (H–x–y) diagram by any line (beg). The points on this line represent the reboiler (x_B, Q_B), the liquid phase of Stage m, $L_m(x_m, h_m)$, and the vapor phase of Stage $m + 1$, $V_{m+1}(y_{n+1}, H_{m+1})$. However, the number of stages in each section can be now obtained by the well-known graphical technique (6).

One of the important features of the Ponchon method is the ability to evaluate the minimum reflux ratio $R_{\min}$. This can be determined by extending the tie line (ff_2) which passes through the feed point (f) to intersect the x_D vertical line at $d_m(x_D, Q_{Dm})$. Q_{Dm} is the minimum heat removed from the condenser. In this case $R_{\min}$ can be estimated by dividing the distance (d_md_2) by the distance (d_2d_1). In other words,

$$R_{\min} = (Q_{Dm} - H_D)/(H_D - h_D) \quad (21)$$

The value of Q_{Dm} is usually obtained from the Ponchon diagram.

THE PROPOSED TECHNIQUE

The proposed technique is essentially based on the features of the Ponchon diagram illustrated in Fig. 2. The minimum reflux ratio can be easily calculated from Eq. (21) if it were possible to estimate the values of Q_{Dm}, h_D, and H_D. However, the enthalpies of distillates h_D and H_D can be estimated from Eqs. (8) and (10) since x_D is specified and y_D is equal to x_D for the case of a total condenser. In the case of a partial condenser,

$y_D = kx_D$. The value of Q_{Dm} can be estimated from the following relation which can be derived from Fig. 2.

$$Q_{Dm} = H_F + (d_m d_3) \tag{22}$$

But

$$(d_m d_3)/(d_3 f_2) = \tan \alpha = (f_2 f_1)/(f_1 f)) = \theta_F \tag{23}$$

Therefore:

$$(d_m d_3) = [(H_F - h_F)/(y_F - x_F)](x_D - y_F) \tag{24}$$

Defining

$$\theta_F = H_F - h_F/y_F - x_F \tag{25}$$

Eq. (22) can be written as

$$Q_{Dm} = H_F + \theta_F(x_D - y_F) \tag{26}$$

The values of y_F, h_F, and H_F can be estimated from polynomials 2, 8, and 10, respectively. Hence Q_{Dm} and $R_{\min}$ can be estimated from Eqs. (26) and (21), respectively.

Some larger reflux ratio, R, must obviously be used for practical separations whereupon the location of Point d should be above Point d_m. This is given by

$$R = aR_{\min} \tag{27}$$

where a is a constant in the range 1.1 to 1.5. The value of R can be used now to obtain the condenser net heat Q_D:

$$R = (Q_D - H_D)/(H_D - h_D) \tag{28}$$

The condenser duty Q_C can then be estimated from Eq. (14). The reboiler net heat Q_B and the reboiler load Q_R can be estimated from Eqs. (13) and (15).

It is now possible to start the computation for the determination of the number of stages in the tower.

For this purpose, Eq. (18), representing the rectifying section, will be written as

$$L_n/D = \theta_n \tag{29}$$

From Eqs. (18) and (29), the values of y_{n+1} and H_{n+1} can be estimated by

$$y_n + 1 = (\theta_n x_n + z_D)/(\theta_n + 1) \tag{30}$$

$$H_{n+1} = (\theta_n h_n + Q_D)/(\theta_n + 1) \tag{31}$$

Similarly, Eq. (20), representing the stripping section, can be written as

$$L_m/D = \theta_m \tag{32}$$

and from Eqs. (20) and (32) it can be proved that

$$y_{m+1} = (\theta_m x_m - x_B)/(\theta_m - 1) \tag{33}$$

$$H_{m+1} = (\theta_m h_m - Q_B)/(\theta_m - 1) \tag{34}$$

Equations from (29) to (34) together with the equilibrium and enthalpy relations (2), (3), (8), and (10) form the basis of the proposed computation technique presented here.

For the rectifying section, for $n = 0$, the condenser plate, Eq. (29) enables one to estimate θ_0,

$$\theta_0 = L_0/D = R$$

Equations (30) and (31) can be used now to estimate y_1 and H_1. Polynomial 3 can be used to get x_1. From Polynomial 4, the value of h_1 corresponding to x_1 can be estimated.

For $n = 1$, θ_1 cannot be estimated unless some initial guessing is introduced. A reasonable initial guess for the value of θ_n can be deduced from Eq. (18). For example:

$$\theta_n^0 = (Q_D - H_n)/(H_n - h_n) \tag{35}$$

Using the value of θ_n^0 in Eqs. (30) and (31) gives values for y_{n+1}^0 and H_{n+1}^0. The value of y_{n+1}^0 is used in Polynomial 6 to get the corresponding value of H_{n+1}. If this value does not agree with that of H_{n+1}^0, a new value θ_n can be obtained from the estimated value of H_{n+1} from Eq. (18). The new value for θ_n is used in Eq. (30) to get a new value for y_{n+1}. The process is interated until convergence within a certain prescribed limit is reached.

The value of x_n obtained is then checked with that of the feed to test the feed stage location in the usual practice. A new value for n is then used until the feed stage location is determined.

The calculation process proceeds into the stripping section using similar steps. An initial guess for the value of θ_m based on Eq. (20) is used where

$$\theta_m^0 = (H_m - Q_B)/(H_m - h_m) \tag{36}$$

Equations (33) and (34) are used in this section together with Polynomials 10 and 8, respectively, to determine the number of equilibrium stages in the bottom section until the reboiler stage concentration is reached.

ALGORITHM

Data Required

i) Antoine constants of the binary system components A_i, B_i, C_i
ii) UNIQUAC equation parameters: q_i, q_i', a_{ij}
iii) Boiling points of the components and operating pressure and temperature
iv) Heat capacities and latent heats of the components
v) Gas constant R
vi) (T–x) data for the system, bubble points t_b and dew points t_d of the mixture
vii) F, $z_{i,F}$, $x_{i,D}$, $x_{i,B}$, a, convergence limits ϵ_1, ϵ_2

Steps

1. *Equilibrium data*
   Use the Antoine equation to get P_{vi}
   Use the UNIQUAC equation to get ϵ_i
   $k_i = \epsilon_i P_{vi}/P$: $y_1 = K_i x_i$
   Correlate $y = P_1(x)$: $x = P_2(y)$
2. *Enthalpy data*
   $\theta_i' = (x_i g_i')/(\Sigma x_i g_i')$: $\tau_{ji} = \exp[-a_{ji}/T]$
   $H_\text{E} = R\Sigma[(q_i' x_i)/\Sigma\theta_j'\tau_{ji})\ (\Sigma\theta_j'\tau_{ji}a_{ji})]$
   $h = \Sigma x_i C_{Li}(t_\text{b} - t_0) + H_\text{E}$
   $H = \Sigma y_i[C_{Li}(t_\text{d} - t_0) + \lambda_i)$
   Correlate $h = P_3(x)$: $H = P_4(y)$
3. *Tower operating conditions*
   Given ($x_{i,D}$, P): estimate the top temperature, T_T, by the dew-point technique
   Given ($x_{i,B}$, P): estimate the bottom temperature, T_B, by the bubble-point technique
   Given ($z_{i,F}$, P), estimate the feed temperature and the composition of the vapor and liquid at equilibrium (y_F, x_F)
4. *Fluids flow rates and enthalpies in the tower*
   $D = F(z_F - x_B)/(x_D - x_B)$: $B = F - D$
   $h_F = P_3(x_F)$: $H_F = P_4(y_F)$
   $h_D = P_3(x_D)$: $H_D = P_4(y_D)$
   $\theta_F = (H_F - h_F)/(y_F - x_F)$
   $Q_{Dm} = H_F + \theta_F(x_D - y_F)$
   $R_\text{min} = (Q_{Dm} - H_D)/(H_D - h_D)$
   $R = aR_\text{min}$
   $Q_D = H_D + R(H_D - h_D)$

$$Q_B = (Fh_F - DQ_D)/B$$
$$Q_C = D(Q_D - h_D)$$
$$Q_R = B(h_B - Q_B)$$

Rectifying and Stripping Sections

```
 1. $n = 0$: $\theta_0 = R$: $x_0 = z_D$ (or $z_D/K$ according to the type of condenser)
    $y_1 = (\theta_0 x_0 + Z_D)/(\theta_0 + 1)$: $H_1 = (\theta_0 h_0 + Q_D)/(\theta_0 + 1)$
    $x_1 = P_2(y_1)$: $h_1 = P_3(x_1)$
 2. $n = 1$
 3. $\theta_n = (Q_0 - H_n)/(H_n - h_n)$
 4. $y^0_{n+1} = (\theta_n x_n + z_D)/(\theta_{n+1})$: $H^0_{n+1} = (\theta_n h_n + Q_D)/(\theta_{n+1})$
    $H_{n+1} = P_4(y^0_{n+1})$
Check: $[(H^0_{n+1} - H_{n+1})/(H^0_{n+1})] \leq \epsilon_1$. Go to 5
    $\theta_n = (Q_D - H_{n+1})/(H_{n+1} - h_n)$. Go to 4
 5. $x_{n+1} = P_2(y_{n+1})$: $h_{n+1} = P_3(x_{n+1})$
Check: $(x_{n+1} - z_F)/(x_{n+1}) \leq \epsilon_2$. Go to 6
    Go to 3
 6. $f_p = n$
    $m = n + 1$
 7. $\theta^0_m = (H_m - Q_B)/(H_m - h_m)$
 8. $y^0_{m+1} = (\theta_m x_m - x_B)/(\theta_m - 1)$: $H^0_{m+1} = (\theta_m h_m - Q_B)/(\theta_m - 1)$
    $H_{m+1} = P_4(y^0_{m+1})$
Check: $[(H^0_{m+1} - H_{m+1})/(H^0_{m+1})] \leq \epsilon_1$. Go to 9
    $\theta_m = (H_{m+1} - Q_B)/(H_{m+1} - h_m)$. Go to 8
 9. $x_{m+1} = P_2(y_{m+1})$: $h_{m+1} = P_3(x_{m+1})$
    Check: $(x_{m+1} - x_B)/(x_{m+1}) \leq \epsilon_2$. Go to 10
    Go to 7
10. Print $D$, $B$, $n$, $f_p$, $m$, $Q_c$, $Q_R$, $x_n$, $y_n$, $h_n$, and $H_n$
```

WORKED EXAMPLE

The developed computation technique was used to solve an example for the distillation of a methanol–water solution.

Treybal (6) has solved this example by the graphical Ponchon–Savarit method. The given data for this problem are:

$F = 216.8$ mol/h
$x_D = 0.915$ mole fraction
$h_f = 1070$ Btu/mol and $R = 1.5R_{\text{min}}$
$x = 0.36$ mole fraction
$x_B = 0.00565$ mole fraction

It is required to calculate $R_{\min}$, N, Q_C, Q_R, L_n, V_n, x_n, y_n, and t_n. The parameters and properties of the binary system are given in Table 1. The equilibrium data were estimated by the UNQUAC equation and compared with the experimental values of Ramalho (7). The maximum deviation in the vapor concentration was about 3%. The enthalpy data were estimated by Eqs. (4), (5), and (9). Agreement between the vapor enthalpies and the published data of Dunlop (8) was good. However, disagreement in the liquid enthalpies was about 9% for some data points. This error may be due to the inaccuracy of the heat of solution estimated by the UNIQUAC technique. The equilibrium and enthalpy data obtained were correlated by fifth-order polynomials. Temperature was also correlated versus x in order to be able to estimate the temperature at each stage in the tower. The polynomial coefficients and the resulting variance for each correlation are given in Table 2.

The computation results are given in Table 3. The number of equilibrium stages obtained was about 7.8 compared with 8.2 obtained by Treybal (6) where the graphical method was used. Stage 4 and not Stage 5 was found to be the feed stage. Differences in the x and y values compared with those obtained by Treybal (6) were 2 to 8%, while in the flow rates, L and V, and the enthalpies, H and h, the differences were 10 to 15%.

Such differences may be attributed to the source of data and also to the relative inaccuracy of the graphical method.

TABLE 1
Parameters and Properties of the Methanol–Water System

Parameter	Methanol, 1	Water, 2	Ref.
Antoine Constants:			
A	9.08097	8.07131	9
B	1582.271	1730.63	
C	239.726	233.426	
UNIQUAC:			
a_{12}	199.3374		9
a_{21}		−109.1007	
r	1.43	0.92	
q	1.43	1.40	
q'	0.96	1.00	
Heat Capacities of Liquids: J/gmol·°C			
C_L at 0–100°C		75.4	10
at 0°C	75.860		
at 20°C	87.072		11
at 40°C	82.590		
Latent Heats, kJ/gmol	33,540	41,698	10

TABLE 2

Equilibrium and Enthalpy Correlations for the System Methanol–Water Data

$y = P_1(x) = 0.034742238 + 5.5460x - 19.65785x^2 + 36.418099x^3 - 33.04203x^4 + 10.711086x^5$

Variance = 0.003225

$x = P_2(y) = 1.7085658\text{E-}3 - 0.2500505y + 3.8561687y^2 - 12.42815y^3 + 17.229539y^4 - 7.418307y^5$

Variance = 0.00208

$h = P_3(x) = 3233.2021 - 33310.31x + 5700.0956x^2 - 5040.732x^3 + 1690.431x^4$

Variance = 24.36

$H = P_4(y) = 20716.57 - 3896.703y + 1668.3591y^2 - 2260.928y^3 + 1163.5129y^4$

Variance = 30.09

$t = P_5(x) = 97.21121 - 101.1154x + 148.7319x^2 - 82.78369x^3$

Variance = 3.4772

OPTIMIZATION

The computation technique can be used for finding the optimum parameters required for a certain system. The developed program was used to find the sensitivity of some parameters to changes in the reflux ratio and the purity of the overhead stream represented by x_D. Table 4 shows the results obtained. The table shows that the number of stages decreases by increasing the reflux ratio, but increases as the required x_D is increased. This is a well-known trend. However, the table also shows that the posi-

TABLE 3

Computational Results

$x_f = 0.3600$ $x_D = 0.9150$ $x_B = 0.00565$ $h_f = 1070.00$

$R_{\min} = 0.79813$ $R = 1.19720$

$Q_{Dm} = 29883.124$ $Q_D = 36008.970$ $Q_B = -21237.431$ $Q_C = 33728.006$ $Q_B = 24451.994$

n	t_n	x_n	y_n	L_n	V_n	H_n	h_n
1	69.1265	0.799800	0.915000	99.19	185.62	17631.43	2328.07
2	71.5691	0.651715	0.852795	96.70	183.65	17819.97	2392.99
3	73.7961	0.479508	0.774530	93.92	181.11	18067.51	2480.17
4	77.3423	0.320941	0.685722	91.69	178.41	18357.25	2602.56
			Feed Plate at $f_n = 4$				
5	81.3594	0.218403	0.606545	322.30	190.18	18619.79	2728.92
6	90.1815	0.078105	0.366802	318.50	189.70	19421.20	3005.47
7	96.0572	0.011610	0.129610	317.48	186.14	20234.95	3195.29
8	97.3467	−0.001337	0.015868	317.34	185.16	20655.15	3237.67

TABLE 4
Optimization of a Distillation Process Using the Simulated Ponchon–Savarit Method
$F = 216.8$, $D = 84.481$, $W = 132.319$,
$x_f = 0.36$, $x_B = 0.00565$, $h_f = 1070$

R	N	f	Q_D	Q_B
(i) $x_D = 0.915$, $R_m = 0.79813$				
$1.1R_m$	9	5	28,827	21,323
$1.3R_m$	8	4	31,277	22,887
$1.5R_m$	8	4	33,728	24,451
(ii) $x_D = 0.95$, $R_m = 0.92852$				
$1.1R_m$	11	7	30,849	21,391
$1.3R_m$	10	6	33,683	23,094
$1.5R_m$	9	5	36,517	24,796
(iii) $x_D = 0.985$, $R_m = 1.06001$				
$1.1R_m$	14	10	32,869	21,452
$1.3R_m$	13	9	36,080	23,276
$1.5R_m$	12	8	39,303	25,100

tion of the feed stage moves up in the tower as the values of the reflux ratio and x_D increase. The table also shows that the reboiler heat load, Q_R, is not very sensitive to changes in the x_D values.

CONCLUSIONS

The Ponchon–Savarit method used in the distillation of binary systems has been simulated. The simulation technique requires that the equilibrium and enthalpy data for the system be accurately estimated and correlated. The technique can be used for calculating and optimizing all the pertinent parameters required. Similar approaches may also be adopted for simulating other stagewise operations.

NOMENCLATURE

a	constant
a_{ij}	interaction parameter
B	bottom, bottom flow rate (mol/h)
C	specific heat (Btu/mol·°F)
D	distillate, distillate flow rate (mol/h)
F	feed flow rate (mol/h)
H	enthalpy of vapor (Btu/mol)

H_E	heat of solution (Btu/mol)
h	enthalpy of liquid (Btu/mol)
i	component
L	liquid flow rate (mol/h)
m	plate number in the stripping section
N	total number of plates in the tower
n	plate number in the rectifying section
P	polynomial, pressure
q, q'	surface area parameters
Q	heat added or removed (Btu/mol)
r	molecular parameter
R	reflux ratio
R_m, R_{min}	minimum reflux ratio
t	temperature (°C or °F)
T	temperature (K or R)
V	vapor flow rate (mol/h)
x	liquid concentration (mole fraction)
y	vapor concentration (mole fraction)
z	feed or distillate concentration (mole fraction)
α	angle
θ	slope of the operating line
θ'	parameter defined by Eq. (6)
λ	latent heat
τ	a UNIQUAC parameter defined by Eq. (7)

Subscripts

b	bubble point
d	dew point

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