Fractional Distillation using Enthalpy-Concentration Method

- molal flow rates need not to be constant
- using enthalpy as well as material balances

When the operating lines are plotted with the equilibrium curve on the x-y diagram, the McCabe-Thiele step-by-step construction can be used to compute the number of ideal plates. However, unless $L_n$ and $L_m$ are constant, the operating lines are curved and can be plotted only if the change with concentration in these internal streams is known. Enthalpy balances are required in the general case to determine the position of a curved operating line.

The enthalpy-concentration diagram requires the following data at a constant pressure:
1. heat capacity of the liquid as a function of temperature, composition, and pressure.
2. heat of solution as a function of temperature and composition.
3. latent heats of vaporization as a function of composition and pressure or temperature.
4. boiling point as a function of pressure, composition and temperature.

The diagram at a given constant pressure is based on arbitrary reference states of liquid and temperature, which is usually taken as the boiling point of the lower-boiling component A.

The saturated liquid enthalpy is
\[ H_x = x_A C_{PLA} (T_b - T_0) + (1 - x_A) C_{PLB} (T_b - T_0) + \Delta H_{sol} \]  (1)

- $T_b$ = boiling temperature of mixture at $x_A$
- $T_0$ = reference temperature
- $C_{PLA}$ = liquid heat capacity of A
- $C_{PLB}$ = liquid heat capacity of B
- $\Delta H_{sol}$ = heat of mixing, usually ignored

The saturated vapor enthalpy is
\[ H_y = y_A [\Delta H_{VA} + C_{PVA} (T_d - T_0)] + (1 - y_A) [\Delta H_{VB} + C_{PVB} (T_d - T_0)] \]  (2)

- $T_d$ = dew temperature of mixture at $y_A$
- $T_0$ = reference temperature
- $C_{PVA}$ = vapor heat capacity of A
- $C_{PVB}$ = vapor heat capacity of B
- $\Delta H_{VA}$ = latent heat of A at the reference temperature $T_0$ & calculated from that at the normal boiling temperature $T_{bA}$.

\[ \Delta H_{VA} = C_{PLA} (T_{bA} - T_0) + \Delta H_{VAb} - C_{PVA} (T_{bA} - T_0) \]  (3)
\[ \Delta H_{VB} = C_{PLB} (T_{bB} - T_0) + \Delta H_{Vbb} - C_{PVB} (T_{bB} - T_0) \]  (4)

That is, the liquid is first changed from temperature $T_0$ to $T_b$ and then evaporated at $T_b$, the vapor temperature is finally changed back from $T_b$ to $T_0$. 
Example D6. Enthalpy-concentration plot for benzene-toluene

Prepare an enthalpy-concentration plot for benzene-toluene at 1 atm pressure. Equilibrium and physical property data are given.

<table>
<thead>
<tr>
<th>T (K)</th>
<th>Vapor pressure (kPa)</th>
<th>Mole fraction of benzene at 101.325 kPa</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Benzene</td>
<td>Toluene</td>
</tr>
<tr>
<td>353.3</td>
<td>101.32</td>
<td>-</td>
</tr>
<tr>
<td>358.2</td>
<td>116.9</td>
<td>46.0</td>
</tr>
<tr>
<td>363.2</td>
<td>135.5</td>
<td>54.0</td>
</tr>
<tr>
<td>368.2</td>
<td>155.7</td>
<td>63.3</td>
</tr>
<tr>
<td>373.2</td>
<td>179.2</td>
<td>74.3</td>
</tr>
<tr>
<td>378.2</td>
<td>204.2</td>
<td>86.0</td>
</tr>
<tr>
<td>383.8</td>
<td>240.0</td>
<td>101.32</td>
</tr>
</tbody>
</table>

Component | Boiling point (°C) | Heat capacity (kJ/kmol K) | Latent heat of vaporization (kJ/kmol)
---|-------------------|-----------------|-------------------|
| Benzene (A) | 80.1 | 138.2 | 96.3 | 30820 |
| Toluene (B) | 110.6 | 167.5 | 138.2 | 33330 |

Solution.

We choose a reference temperature of $T_0=80.1^\circ C$ so that the liquid enthalpy of pure benzene ($x_A=1$) at the boiling point is zero.

For the first point, we select pure toluene ($x_A=0$). For liquid toluene at the boiling point of 110.6°C,

$$H_x = x_A C_{PLA}(T_b - T_0) + (1 - x_A) C_{PLB}(T_b - T_0) + \Delta H_{sol}$$

$$= 0 + (1-0)(167.5)(110.6 - 80.1) = 5109 \text{ kJ/kmol}$$

For the saturated vapor enthalpy of pure toluene, we first need to calculate $\Delta H_{VB}$ at $T_0=80.1^\circ C$:

$$\Delta H_{VB} = C_{PLB}(T_b - T_0) + \Delta H_{VB} - C_{PVB}(T_b - T_0)$$

$$= 167.5(110.6 - 80.1) + 33330 - 138.2(110.6 - 80.1)$$

$$= 34224 \text{ kJ/kmol}$$

Hence, at $x_A = 0$,

$$H_y = y_A[\Delta H_{VA} + C_{PV}(T_d - T_0)] + (1 - y_A)[\Delta H_{VB} + C_{PVB}(T_d - T_0)]$$

$$= 0 + (1 - 0)\left[34224 + 138.2(110.6 - 80.1]\right]$$

$$= 38439 \text{ kJ/kmol}$$

For pure benzene, $x_A = y_A = 1$, $T = T_0=80.1^\circ C$, $H_x = 0$,

$$H_y = y_A[\Delta H_{VA} + C_{PV}(T_d - T_0)] + (1 - y_A)[\Delta H_{VB} + C_{PVB}(T_d - T_0)]$$

$$= 1\left[30820 + 96.3(80.1 - 80.1)\right] + 0 = 30820 \text{ kJ/kmol}$$

Selecting $x_A = 0.5$, the boiling point $T_b = 92^\circ C$ and the dew point for $y_A = 0.5$ is $T_d = 98.8^\circ C$ from the T-x-y plot.

$$H_x = x_A C_{PLA}(T_b - T_0) + (1 - x_A) C_{PLB}(T_b - T_0)$$

$$= 0.5(138.2)(92-80.1) + (1-0.5)(167.5)(92-80.1)$$

$$= 1820 \text{ kJ/kmol}$$

$$H_y = y_A[\Delta H_{VA} + C_{PV}(T_d - T_0)] + (1 - y_A)[\Delta H_{VB} + C_{PVB}(T_d - T_0)]$$

$$= 0.5\left[30820 + 96.3(98.8 - 80.1)\right] + (1 - 0.5)\left[34224 + 138.2(98.8 - 80.1)\right]$$

$$= 34176 \text{ kJ/kmol}$$
By the similar procedure, at $x_A = 0.3$ and $y_A = 0.3$,
- $H_x = 2920 \text{ kJ/kmol}$ and $H_y = 36268 \text{ kJ/kmol}$

at $x_A = 0.8$ and $y_A = 0.8$,
- $H_x = 562 \text{ kJ/kmol}$ and $H_y = 32380 \text{ kJ/kmol}$

Enthalpy-concentration data for benzene-toluene mixture at 1 atm.

<table>
<thead>
<tr>
<th>Saturated Liquid</th>
<th>Saturated Vapor</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mole fraction, $x_A$</td>
<td>Enthalpy, $H_x$, (kJ/kmol)</td>
</tr>
<tr>
<td>0.30</td>
<td>2920</td>
</tr>
<tr>
<td>0.50</td>
<td>1820</td>
</tr>
<tr>
<td>0.80</td>
<td>562</td>
</tr>
<tr>
<td>1.00</td>
<td>0</td>
</tr>
</tbody>
</table>

Enthalpy-concentration plot for benzene-toluene mixture at 1 atm. The tie line represents the enthalpies and composition of the liquid and vapor phases in equilibrium.

**Distillation in Enriching Section of Tower**

Material balances

\[
V_{n+1} = L_n + D     \quad (5)
\]

\[
V_{n+1}y_{n+1} = L_n x_n + D x_D     \quad (6)
\]

By rearranging Eq. (6) we obtain the operating line

\[
y_{n+1} = \frac{L_n}{V_{n+1}} x_n + \frac{D x_D}{V_{n+1}}     \quad (7)
\]

$L_n$ and $V_{n+1}$ may vary throughout the tower so Eq. (7) will not be a straight line on an $xy$ plot.

Making an enthalpy balance,

\[
V_{n+1} H_{y,n+1} = L_n H_{x,n} + D H_{x,D} + q_c \quad (8)
\]

where $q_c$ is the condenser duty, kJ/h.

An enthalpy balance for the condenser only gives

\[
q_c = V_1 H_{y,1} - L H_{x,D} - D H_{x,D} \quad (9)
\]
Substitute Eq. (9) into Eq. (8) we have

\[
V_{n+1}H_{y,n+1} = L_n H_{xn} + V_1 H_{y1} - LH_{xD}
\]

\[
= (V_{n+1} - D)H_{xn} + V_1 H_{y1} - LH_{xD}
\]

Eqs. (7) & (10) are the final working equations for the enriching section.

In order to plot the operating line Eq. (7), the terms \(V_{n+1}\) and \(L_n\) must be determined from Eq. (10). If the reflux ratio is set, \(V_1\) and \(L\) are known. The values of \(H_y1\) and \(H_{xD}\) can be determined by Eqs. (1) & (2) or from an enthalpy-concentration plot. If a value of \(x_n\) is selected, it is a trial-and-error solution to obtain \(H_{y,n+1}\) since \(y_{n+1}\) is unknown. The steps to follow are given below.

1. Select a value of \(x_n\). Assume \(V_{n+1} = V_1 = L + D\), then \(L_n = V_{n+1} - D\).
2. Use Eq. (7) to calculate the approximate value of \(y_{n+1}\).
3. Using this \(y_{n+1}\) to obtain \(H_{y,n+1}\) and \(x_n\) to obtain \(H_{xn}\). Substitute these values into Eq. (10) and solve for the new \(V_{n+1}\). Obtain new \(L_n\) from Eq. (5).
4. Substitute the above values into Eq. (7) to get the new \(y_{n+1}\).
5. If the newly calculated \(y_{n+1} \neq \) the assumed value, repeat steps 3-4. Usually one iteration is enough. Assume another \(x_n\) and repeat steps 2-5.
6. Plot the curved operating line for the enriching section.

**Distillation in Stripping Section of Tower**

Material balances

\[
L_m = V_{m+1} + B
\]

\[
L_m x_m = V_{m+1} y_{m+1} + B x_B
\]

so that the operating line is

\[
y_{m+1} = \frac{L_m}{V_{m+1}} x_m - \frac{B x_B}{V_{m+1}}
\]

Making an enthalpy balance,

\[
V_{m+1} H_{y,m+1} = (V_{m+1} + B) H_{xm} - B H_{xB} + q_r
\]

where \(q_r\) is the reboiler duty, kJ/h.

An enthalpy balance for the whole distiller gives

\[
q_r = D H_{xD} + B H_{xB} + q_e - F H_F
\]

The final working equations are Eqs. (13) and (15).

Using a method similar to that of the enriching section to solve the equations.

<table>
<thead>
<tr>
<th>Component</th>
<th>(T_b) (°C)</th>
<th>(C_p) (kJ/(kmol K))</th>
<th>(\Delta H_{vb}) (kJ/kmol)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Benzene</td>
<td>80.1</td>
<td>Liquid 138.2</td>
<td>Vapor 96.3</td>
</tr>
<tr>
<td>Toluene</td>
<td>110.6</td>
<td>Liquid 167.5</td>
<td>Vapor 138.2</td>
</tr>
</tbody>
</table>
Example D7. Distillation Using Enthalpy-Concentration Method

A liquid mixture of benzene-toluene is to be distilled in a fractionating tower at 101.3 kPa pressure. The feed of 100 kmol/h is liquid, containing 45 mol % benzene and 55 mol % toluene, and enters at 54.4°C. A distillate containing 95 mol % benzene and 5 mol % toluene and a bottoms containing 10 mol % benzene and 90 mol % toluene are to be obtained. The reflux ratio is 1.755. The average heat capacity of the feed is 159 kJ/(kmol K) and the average latent heat is 32099 kJ/kmol. Use enthalpy balances to calculate the flow rates of the liquid and vapor at various points in the tower and plot the curved operating lines. Determine the number of theoretical stages needed.

Solution:

\[ F = 100 \text{ kg mol/h}, \ x_F = 0.45, \ x_D = 0.95, \ x_B = 0.10, \]
\[ R = 1.5R_m = 1.5(1.17) = 1.755, \]
\[ D = 41.2 \text{ kg mol/h}, \ B = 58.8 \text{ kmol/h} \]

The feed enters at 54.5°C so \( q = 1.195 \).

The flows at the top of the tower are calculated as follows.

\[ R = L/D = 1.755; \ L = 1.755D = 1.755(41.2) = 72.3; \]
\[ V_1 = L+D = 72.3+41.2 = 113.5 \]

The latent heats of vaporization of benzene and toluene at the reference temperature of 80.1°C are

\[ \Delta H_{VA} = \Delta H_{VAb} = 30820 \text{ kJ/kmol} \]

\[ \Delta H_{VB} = C_{PLB}(T_{bB} - T_0) + \Delta H_{VBB} - C_{PVB}(T_{bB} - T_0) \]
\[ = 167.5(110.6-80.1) + 33330 - 138.2(110.6-80.1) \]
\[ = 34224 \text{ kJ/kmol} \]

The saturation temperature (dew point) at the top of the tower for \( y_1 = x_D = 0.95 \) is 82.3°C.

\[ H_{y_1} = 0.95[30820+96.3(82.3-80.1)] \]
\[ + (1-0.95)[34224+138.2(82.3-80.1)] = 31206 \text{ kJ/kmol} \]

The boiling point of the distillate \( D \) is obtained as 81.1°C.

\[ H_{xD} = 0.95(138.2)(81.1-80.1) \]
\[ + (1-0.95)(167.5)(81.1-80.1) = 139 \text{ kJ/kmol} \]

Step 1: select \( x_n = 0.55 \) and guess \( y_{n+1} \) from Eq. (7)

\[ y_{n+1} = \frac{L_n}{V_{n+1}} x_n + \frac{Dx_D}{V_{n+1}} = \frac{72.3}{113.5}(0.55) + \frac{41.2}{113.5}(0.95) = 0.695 \]

Step 2: using the Figure in Example D6 for \( x_n = 0.55 \) obtain \( H_{xn} = 1590 \) and for \( y_{n+1} = 0.695, H_{y,n+1} = 33240 \).

Substituting into Eq. (10)

\[ V_{n+1} (33240) = (V_{n+1} - 41.2)1590 \]
\[ +113.5(31206) - 72.3(139) \]
\[ V_{n+1} = 109.5 \]

Using Eq. (5),

\[ 109.5 = L_n + 41.2 \] or \( L_n = 68.3 \)

Step 3: substituting into Eq. (7),

\[ y_{n+1} = \frac{L_n}{V_{n+1}}(0.55) + \frac{Dx_D}{V_{n+1}}(0.95) = 0.700 \]
This calculated value is close enough to the approximate value of 0.695 so that no further trials are needed.

Select other values of \(x_n\) and calculate the responding \(y_{n+1}\), plot the curved operating line in the enriching section.

The condenser duty is obtained from Eq. (9)

\[
q_c = V_f H_f + L H_{xD} - DH_{xD}
\]

\[
= 113.5(31206) - 72.3(139) - 41.2(139) = 3526100 \text{ kJ/h}
\]

For \(x_B = 0.10\), \(H_{xB} = 4350\) from the Figure in page 89.

The feed is at 54.5\(^\circ\)C, using Eq. (1), we have

\[H_F = 0.45(138.2)(54.5-80.1) + (1-0.45)(167.5)(54.5-80.1) = -3929 \text{ kJ/kg mol}\]

Using Eq. (15),

\[q_r = 41.2(139) + 58.8(4350) + 3526100 - 100(-3929) = 4180500 \text{ kJ/h}\]

Making a material balance below the bottom tray and around the reboiler we have

\[L_N = B + V_b\] (16)

Enthalpy balance

\[V_b H_{yb} = (V_b + B) H_{xN} - BH_{xB} + q_r\] (17)

From the equilibrium diagram we find that for \(x_B = 0.10\), \(y_B = 0.207\), which is the vapor composition leaving the reboiler.

For the equimolal overflow in the stripping section,

\[L_m = L_n + qF = 72.3 + 1.195(100) = 191.8\] (18)

\[V_{m+1} = V_{n+1} -(1-q)F = 113.5-(1-1.195)100 = 133.0\]

Selecting \(y_{m+1} = y_B = 0.207\) and using Eq. (13), an approximate value of \(x_m = x_N\) is obtained.

\[y_{m+1} = \frac{L_m}{V_{m+1}} x_m - \frac{Bx_B}{V_{m+1}}\]

\[0.207 = \frac{191.8}{133.0} x_N - \frac{58.8}{133.0} (0.10)\]

Solving \(x_N = 0.174\). From the Figure in example D6 for \(x_N = 0.174\), \(H_{xN} = 3800\), and for \(y_B = 0.207\), \(H_{yb} = 37000\).

Substituting into Eq. (17),

\[V_b(37000) = (V_b + 58.8)3800 - 58.8(4350) + 4180500\]

Solving \(V_b = 125.0\). Using Eq. (16) we get \(L_N = 183.8\).

Substituting into Eq. (13) and solving for \(x_N\).

\[0.207 = \frac{183.8}{125.0} x_N - \frac{58.8}{125.0} (0.10)\]

\[x_N = 0.173\]

This value is quite close to the approximate value of 0.174.

Assuming other values of \(y_{m+1}\) and solving for \(x_m\) using the same procedure, the curved operating line for the stripping section can be constructed.

The number of theoretical stages is determined to be 10.4, compared with 9.9 steps using the constant flow method.
Ponchon and Savarit method

Rectifying section

We rearrange Eq. (8) as

\[ V_{n+1}H_{y,n+1} - L_nH_{x_n} = DH_{xD} + q_c = DQ' \]  \hspace{1cm} (16)

where \[ Q' = \frac{DH_{xD} + q_c}{D} \]

The left hand side is the net rate flow of heat upwards through the enriching section, while the right hand side is constant. This means that the net rate of flow of heat is constant and independent of the tray number in the rectifying section.

Using Eq. (5) to eliminate \( V_{n+1} \) in Eq. (6), we have

\[ (L_n + D)y_{n+1} - L_nx_n = DxD \]
\[ L_n(y_{n+1} - x_n) = D(x_D - y_{n+1}) \]  \hspace{1cm} (17)

Hence,

\[ \frac{L_n}{D} = \frac{(x_D - y_{n+1})}{(y_{n+1} - x_n)} = \text{internal reflux ratio} \]  \hspace{1cm} (18)
This can be explained by the lever rule. A vapor of \( V_{n+1} \) of composition \( y_{n+1} \) is separated into a distillate \( D \) of \( x_D \), and a liquid \( L_n \) of \( x_n \). \((V_{n+1}, y_{n+1})\) is at the fulcrum of a beam, balanced by the distillate \((D, x_D)\) and liquid \((L_n, x_n)\). In order for the beam to be “balance”, the moments about the fulcrum must be equal. Hence, we have Eq. (17).

A similar analysis for Eq. (16) gives
\[
(L_n + D)H_{y,n+1} - L_nH_{xn} = DQ'
\]
\[
L_n(H_{y,n+1} - H_{xn}) = D(Q' - H_{y,n+1})
\]

or
\[
\frac{L_n}{D} = \frac{(Q' - H_{y,n+1})}{(H_{y,n+1} - H_{xn})} = \text{internal reflux ratio}
\]

Again this obeys the lever rule.

\[
\frac{L_n}{D} = \frac{(x_D - y_{n+1})}{(y_{n+1} - x_n)} = \frac{(Q' - H_{y,n+1})}{(H_{y,n+1} - H_{xn})}
\]

Combining Eqs. (18) & (19), we have

\[
R = \frac{L_0}{D} = \frac{(x_D - y_1)}{(y_1 - x_D)} = \frac{(Q' - H_{y,1})}{(H_{y,1} - H_{xD})} = \frac{\text{line length } Q' - H_{y,1}}{\text{line length } H_{y,1} - H_{xD}}
\]

Since the heat and mass balances apply to any tray, then at the top plate, \( n = 0 \), we have

\[
R = \frac{L_0}{D} = \frac{(x_D - y_1)}{(y_1 - x_D)} = \frac{(Q' - H_{y,1})}{(H_{y,1} - H_{xD})} = \frac{\text{line length } Q' - H_{y,1}}{\text{line length } H_{y,1} - H_{xD}}
\]

The line length \( H_{y,1} - H_{xD} \) is the latent heat of vaporization at a distillate composition \( x_D \). If the reflux ratio is specified, then the length of \( Q' - H_{y,1} \) can be calculated, and hence \( Q' \) determined.
Equilibrium data are used to determine the liquid and vapor compositions leaving the first theoretical plate. The graphical solution of Ponchon and Savarit is continued using the equilibrium data and the H-x-y diagram until the reboiler composition is reached. In the following case, a total of 4 theoretical plates is shown, 3 in the column and 1 for the reboiler having an enthalpy $H_R$ and composition $x_R$.

**Stripping section**

The same analytical techniques used in the rectifying section can be applied to the stripping section and the entire column.

We rearrange Eq. (14) as

$$L_m H_{x,m} - V_{m+1} H_{y,m+1} = B H_{x,B} - q_r = B Q''$$

where $Q'' = H_{x,B} - q_r B$, by eliminating $B (B = L_m - V_{m+1})$,

$$\frac{L_m}{V_{m+1}} = \frac{(y_{m+1} - x_B)}{(x_B - x_m)} = \frac{(H_{y,m+1} - Q'')}{(H_{x,m} - Q'')} \quad (21)$$

On the H-x-y diagram, this is a straight line passing through $(H_{y,m+1}, y_{m+1})$ at $V_{m+1}$, $(H_{x,m}, x_m)$ at $L_m$, and $(Q'', x_B)$ at $\Delta_B$ as shown in the figure above. $\Delta_B$ is also termed the difference point. $B$ is $W$ in the figure.

Eq. (21) applies to every plate in the stripping section. On the H-x-y diagram, this is a line from $V_{Z+1}$ (vapor leaving the reboiler and entering the bottom tray $Z$ of the
tower) to $\Delta_B$, intersects the saturated liquid enthalpy curve at $L_Z$, the liquid leaving the bottom plate. Vapor $V_Z$ leaving the bottom plate is in equilibrium with the liquid $L_Z$ and is located on the tie line $Z$.

The number of theoretical plates in the stripping section can be determined from the H-x-y diagram by alternatively constructing lines to $\Delta_B$ and tie lines. Each tie line is a theoretical plate.

The complete column

The overall mass balance is

$$F = D + W$$

The more volatile component mass balance is

$$Fx_F = DX_D + WX_W$$

The heat balance is

$$FH_F = QC + DH_{x_D} + WH_{x_W} - QR = DQ' + WQ''$$

Eliminating $F$ yields

$$\frac{D}{W} = \frac{(x_F - x_W)}{(x_D - x_F)} = \frac{(H_F - Q'')}{(Q' - H_F)}$$

This is a straight line on the H-x-y diagram passing through $(Q', x_D)$ at $\Delta_D$, $(H_F, x_F)$ at F, and $(Q'', x_W)$ at $\Delta_W$ and is shown in the above figure for a case of liquid feed below its boiling point (cold liquid).

The procedure to determine the number of theoretical plates is summarized below.

1. Locate the feed enthalpy and composition
2. Locate $x_D$ and $x_W$
3. Locate $\Delta_D$ by computation of $Q'$ or for a specific reflux ratio as $R = L_0/D$
4. The line $\Delta_D$, F is extended to $x = x_W$ and thus locates $\Delta_W$ whose coordinates can be used to calculate $Q_R$.
5. The number of theoretical plates is then determined by application of the straight line relationship on the H-x-y diagram and the equilibrium data of the y-x diagram.
Example D8. Distillation Using Ponchon Savarit Method

Re-calculate the number of theoretical stages required for example D7 using Ponchon Savarit method.

\[ x_F = 0.45, \quad H_F = -3929 \text{ kJ/kg mol} \]
\[ x_D = 0.95, \quad x_B = 0.10 \]
\[ Q' = \frac{D H_{xD} + q_c}{D} = \frac{41.2(139) + 3526100}{41.2} = 85724 \text{ kJ/kmol} \]

\[ Q'' = H_{xB} - q_r/B = 435 -4180500/58.8 = -66747 \text{ kJ/kmol} \]

or in the H-x-y diagram, draw the line from \( \Delta_D \) \((x_D=0.95, \quad Q' = 85724)\) through \((x_F=0.45, \quad H_F = -3929)\) to intersect \( x_B = 0.10 \) gives \( Q'' = -66747 \text{ kJ/kmol} \) from \( \Delta_B \).

The number of theoretical plates is then determined by application of the straight line relationship on the H-x-y diagram and the equilibrium data of the y-x diagram. 10.4 plates, or 9.4 plates plus 1 reboiler are needed. The feed is at plate 6.