

MODULE 5: DISTILLATION

LECTURE NO. 5

Determination of the stripping section operating line (SOL):

The stripping section operating line (SOL) can be obtained from the ROL and q -line without doing any material balance. The SOL can be drawn by connecting point x_B on the diagonal to the point of intersection between the ROL and q -line. The SOL will change if q -line is changed at fixed ROL. The change of SOL with different q -lines for a given ROL at constant x_D is shown in Figure 5.16.

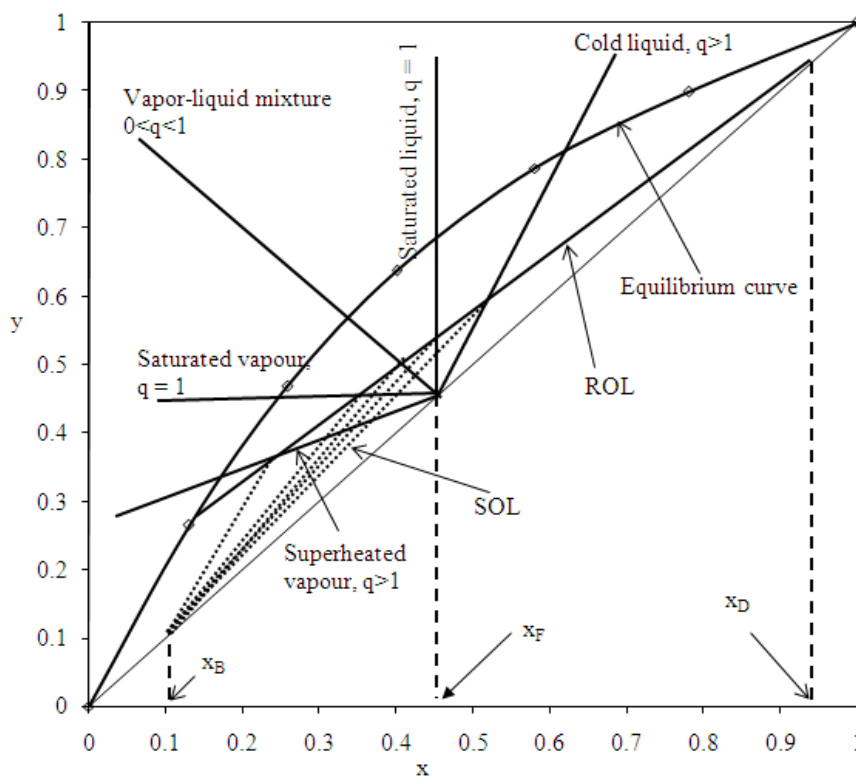


Figure 5.16: Stripping section operating line with different q -lines

The stripping section operating line can be derived from the material balance around the stripping section of the distillation column. The stripping section of a distillation column is shown in Figure 5.17. The reboiled vapor is in equilibrium with bottoms liquid which is leaving the column.

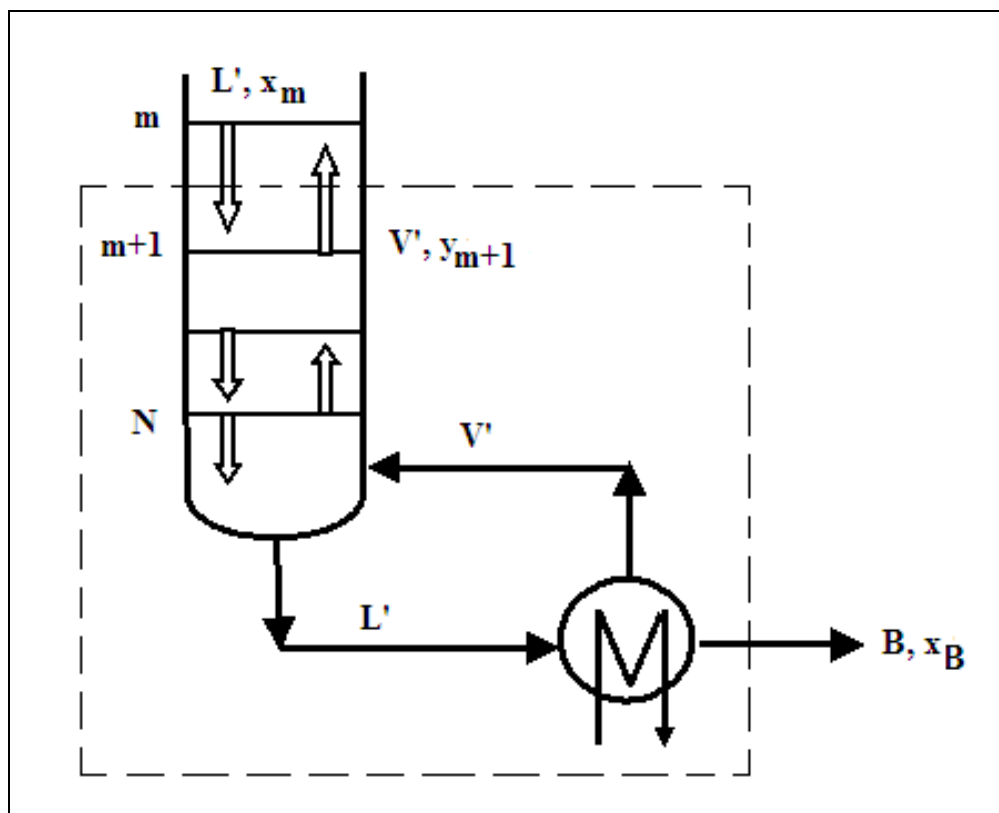


Figure 5.17: Schematic of the stripping section

Consider the constant molal overflow in the column. Thus $L'_m = L'_{m+1} = \dots = L' =$ constant and $V'_m = V'_{m+1} = \dots = V' =$ constant.

Overall material balance gives

$$L' = V' + B \quad (5.29)$$

More volatile component balance gives:

$$L'x_m = V'y_{m+1} + Bx_B \quad (5.30)$$

Substituting and re-arranging the Equation (5.30) yields

$$y_{m+1} = \left(\frac{L'}{V'} \right) x_m - \left(\frac{B}{V'} \right) x_B \quad (5.31)$$

Dropping the subscripts "m+1" and "m" it becomes:

$$y = \left(\frac{L'}{V'} \right) x - \left(\frac{B}{V'} \right) x_B \quad (5.32)$$

Substituting $V' = L' - B$ from Equation (5.29), the Equation (5.32) can be written as:

$$y = \left(\frac{L'}{L' - B} \right) x - \left(\frac{B}{L' - B} \right) x_B \quad (5.33)$$

The Equation (5.33) is called the stripping operating line (SOL) which is a straight line with slope $(L' / L' - B)$ and intercept $(B x_B / L' - B)$. When $x = x_B$, $y = x_B$, the SOL passes through (x_B, x_B) on the 45° diagonal line.

Determination of number of theoretical stage

Suppose a column is to be designed for the separation of a binary mixture where the feed has a concentration of x_F (mole fraction) of the more volatile component and a distillate having a concentration of x_D of the more volatile component whereas the bottoms having a desired concentration of x_B . Once the three lines (ROL, SOL and q-line) are drawn, the number of theoretical stages required for a given separation is then the number of triangles that can be drawn between these operating lines and the equilibrium curve. The last triangle on the diagram represents the reboiler. A typical representation is given in Figure 5.18.

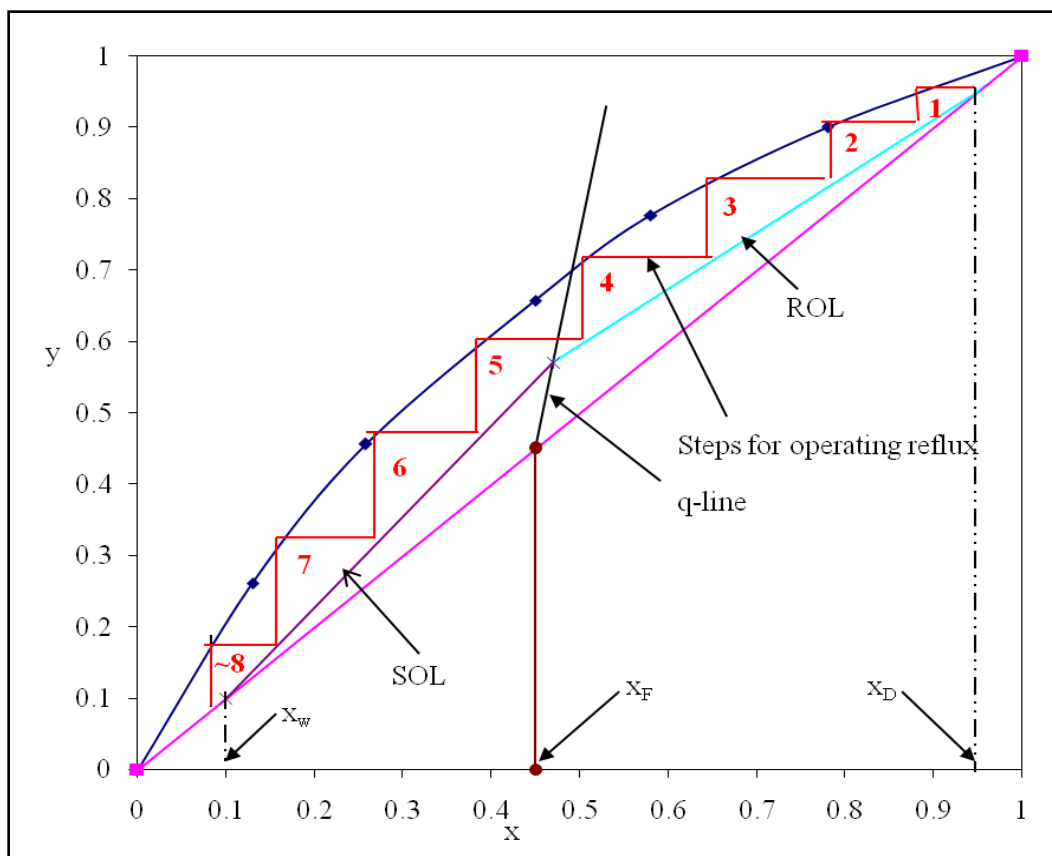


Figure 5.18: A typical representation of identifying number of theoretical stages

Reflux Ratio, R

The separation efficiency by distillation depends on the reflux ratio. For a given separation (i.e. constant x_D and x_B) from a given feed condition (x_F and q), higher reflux ratio (R) results in lesser number of required theoretical trays (N) and vice versa. So there is an inverse relationship between the reflux ratio and the number of theoretical stages. At a specified distillate concentration, x_D , when R changes, the slope and intercept of the ROL changes (Equation (5.19)). From the Equation (5.19), when R increases (with x_D constant), the slope of ROL becomes steeper, i.e. $(R/R+1)$ and the intercept $(x_D/R+1)$ decreases. The ROL therefore rotates around the point (x_D, x_D) . The reflux ratio may be any value between a minimum value and an infinite value. The limit is the minimum reflux ratio (result in infinite stages) and the total reflux or infinite reflux ratio (result in minimum stages). With x_D constant, as R decreases, the slope $(R/R+1)$ of ROL (Equation

(5.19)) decreases, while its intercept ($x_D/R+1$) increases and rotates upwards around (x_D, x_D) as shown in Figure 5.19. The ROL moves closer to the equilibrium curve as R decreases until point Q is reached. Point Q is the point of intersection between the q-line and the equilibrium curve.

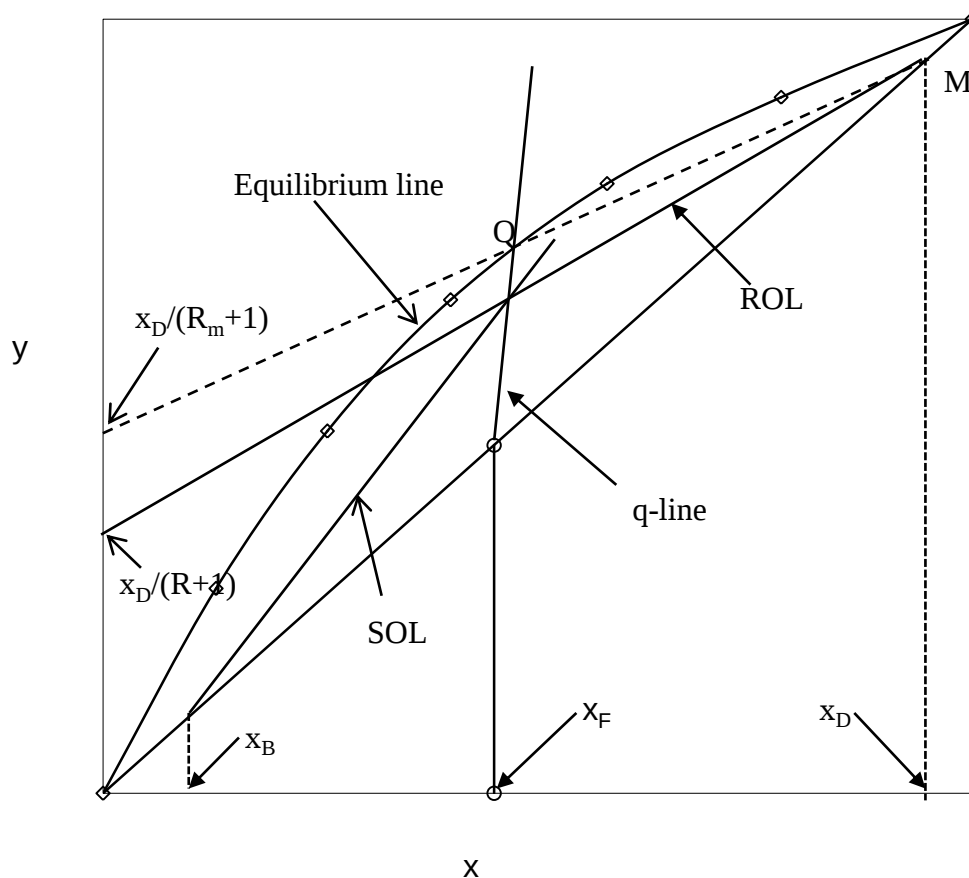


Figure 5.19: Representation of minimum reflux ratio

At this point of intersection the driving force for mass transfer is zero. This is also called as Pinch Point. At this point separation is not possible. The R cannot be reduced beyond this point. The value of R at this point is known as the minimum reflux ratio and is denoted by $R_{\min}$. For non-ideal mixture it is quite common to exhibit inflections in their equilibrium curves as shown in Figure 5.20 (a, b). In those cases, the operating lines where it becomes tangent to the equilibrium curve (called tangent pinch) is the condition for minimum reflux. The ROL cannot

move beyond point P, e.g. to point K. The condition for zero driving force first occurs at point P, before point K which is the intersection point between the q-line and equilibrium curve. Similarly it is the condition SOL also. At the total reflux ratio, the ROL and SOL coincide with the 45 degree diagonal line. At this condition, total number of triangles formed with the equilibrium curve is equal minimum number of theoretical stages. The reflux ratio will be infinite.

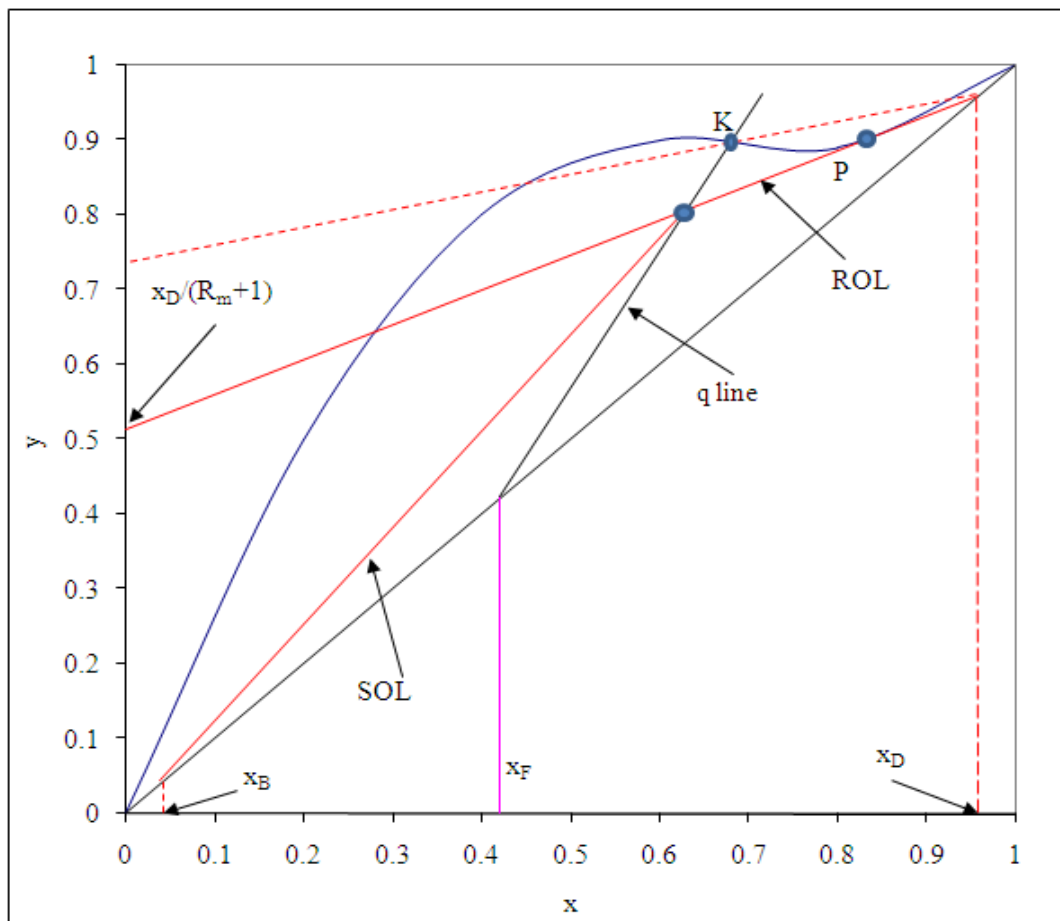


Figure 5.20 (a): Representation of minimum reflux ratio for non-ideal mixture

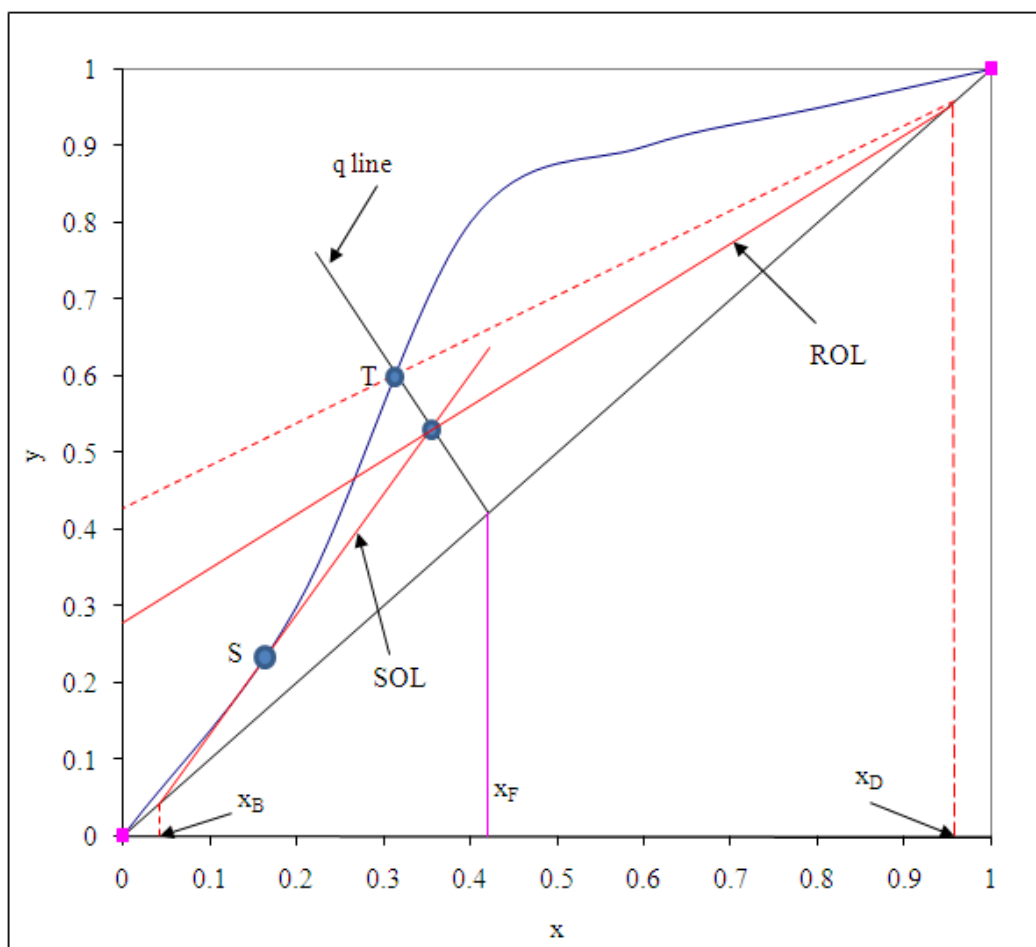


Figure 5.20 (b): Representation of minimum reflux ratio for non-ideal mixture

Tray Efficiency

For the analysis of theoretical stage required for the distillation, it is assumed that the the vapor leaving each tray is in equilibrium with the liquid leaving the same tray and the trays are operating at 100% efficiency. In practice, the trays are not perfect. There are deviations from ideal conditions. The equilibrium with temperature is sometimes reasonable for exothermic chemical reaction but the equilibrium with respect to mass transfer is not often valid. The deviation from the ideal condition is due to: (1) Insufficient time of contact (2) Insufficient degree of mixing. To achieve the same degree of desired separation, more trays will have to be added to compensate for the lack of

perfect separability. The concept of tray efficiency may be used to adjust the the actual number of trays required.

Overall Efficiency

The overall tray efficiency, E_o is defined as:

$$E_o = \frac{\text{No. of theoretical trays}}{\text{No. of actual trays}} \quad (5.34)$$

It is applied for the whole column. Every tray is assumed to have the same efficiency. The overall efficiency depends on the (i) geometry and design of the contacting trays, (ii) flow rates and flow paths of vapor and liquid streams, (iii) Compositions and properties of vapor and liquid streams (Treybal, 1981; Seader and Henley, 1998). The overall efficiency can be calculated from the following correlations:

The Drickamer-Bradford empirical correlation:

$$E_o = 13.3 - 66.8 \log(\mu) \quad (5.35)$$

The correlation is valid for hydrocarbon mixtures in the range of $342 \text{ K} < T < 488.5 \text{ K}$, $1 \text{ atm} < P < 25 \text{ atm}$ and $0.066 < \mu < 0.355 \text{ cP}$

The O'Connell correlation:

$$E_o = 50.3(\alpha\mu)^{-0.226} \quad (5.36)$$

Murphree Efficiency

The efficiency of the tray can also be calculated based on semi-theoretical models which can be interpreted by the Murphree Tray Efficiency E_M . In this case it is assumed that the vapor and liquid between trays are well-mixed and have uniform composition. It is defined for each tray according to the separation

achieved on each tray based on either the liquid phase or the vapor phase. For a given component, it can be expressed as:

Based on vapor phase:

$$E_{MV} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} \quad (5.37)$$

Based on liquid phase:

$$E_{ML} = \frac{x_n - x_{n-1}}{x_n^* - x_{n-1}} \quad (5.38)$$

Example problem 5.2:

A liquid mixture of benzene toluene is being distilled in a fractionating column at 101.3 k Pa pressure. The feed of 100 kmole/h is liquid and it contains 45 mole% benzene (A) and 55 mole% toluene (B) and enters at 327.6 K. A distillate containing 95 mole% benzene and 5 mole% toluene and a bottoms containing 10 mole% benzene and 90 mole% toluene are to be obtained. The amount of liquid is fed back to the column at the top is 4 times the distillate product. The average heat capacity of the feed is 159 KJ/kg mole. K and the average latent heat 32099 kJ/kg moles.

Calculate

- i. The kg moles per hour distillate, kg mole per hour bottoms
- ii. No. of theoretical stages at the operating reflux.
- iii. The minimum no. of theoretical stages required at total reflux
- iv. If the actual no. of stage is 10, what is the overall efficiency increased at operating condition compared to the condition of total reflux?

The equilibrium data:

Temp.(K)	353.3	358.2	363.2	366.7	373.2	378.2	383.8
x_A (mole fraction)	1.000	0.780	0.580	0.450	0.258	0.13	0
y_A (mole fraction)	1.000	0.900	0.777	0.657	0.456	0.261	0

Solution 5.2:

$$F = D + B$$

$$100 = D + B$$

$$F x_F = D x_D + B x_B$$

Therefore, $D = 41.2$ kg mole/h, $B = 58.8$ kg mole/h

$$y = [R/(R+1)] x + x_D/(R+1) = 0.8 x + 0.190$$

$$q = 1 + cpL (T_B - T_F)/\text{Latent heat of vaporization}$$

$T_B = 366.7$ K from boiling point of feed, $T_F = 327.6$ K (inlet feed temp)

Therefore $q = 1.195$

Slope of q line = 6.12

From the graph (Figure E1), Total no of theoretical stages is 8 at operating reflux (Red color)

From the graph (Figure E1), Total no of theoretical stages is 6 at total reflux (Black color)

Overall efficiency at operating conditions: E_o (Operating) = No of ideal stage/ No of actual trays = $7.9/10 = 0.79$

Overall efficiency at total reflux conditions: E_o (total reflux) = No of ideal stage/ No of actual trays = $5.9/10 = 0.59$

Overall efficiency increased: $0.79 - 0.59 = 0.20$

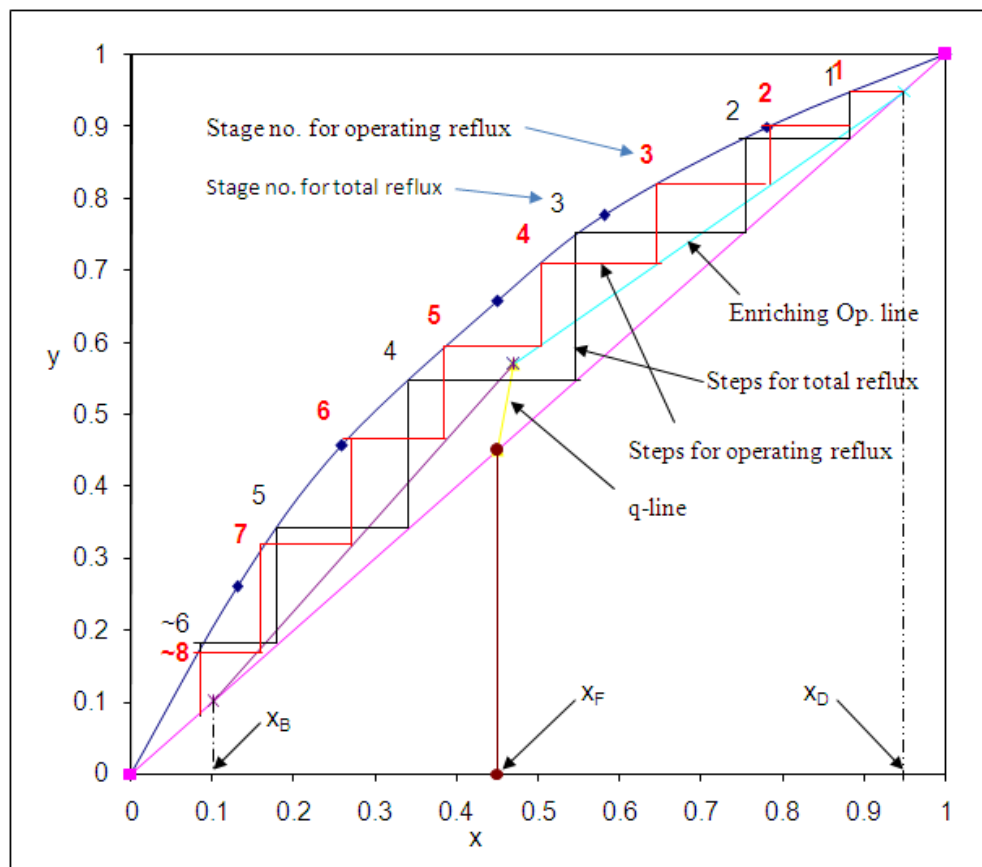


Figure E1: Graph representing the example problem 5.2.