## Continuous Fractionation (Rectification):

### Fractionation Process:

There are three methods used in distillation, each offering a varying degree of separation:

- Flash or equilibrium distillation
- Differential distillation
- Continuous Fractionation

Fractionation is the most important as it offers a greater degree of separation. It is the most widely used unit operation process in the chemical industry. It is a multistage counter-current operation, where the liquid flows counter current to vapor with the feed introduced at some intermediate point. Some of the vapor is condensed (mainly lvc) and some of the liquid is vaporized (mainly mvc). This partial condensation and vaporization continuously enriches the vapor phase with the mvc, with the net result of a better separation.

The operation takes place in a vertical cylindrical pressure vessel (called distillation column or tower, or a fractionator), divided into compartments by a series of perforated plates which permit the upward flow of vapor. The fractionator consists of two main sections:

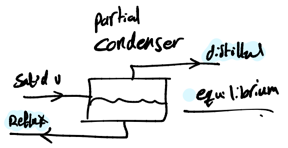
- Top section above the feed point called absorption, enriching, or rectifying section
- Bottom section below the feed point called exhausting or stripping section.

The feed is introduced continuously into the fractionator where it is eventually split into two products. One removed from the top of the column richer than the feed in the mvc (called distillate, top product, overheads product, or light product); the other product withdrawn from the bottom of the column weaker in the mvc (called bottom product, waste, residue or heavy product). The purities of the products depend on the liquid/gas ratio and the number of stages.

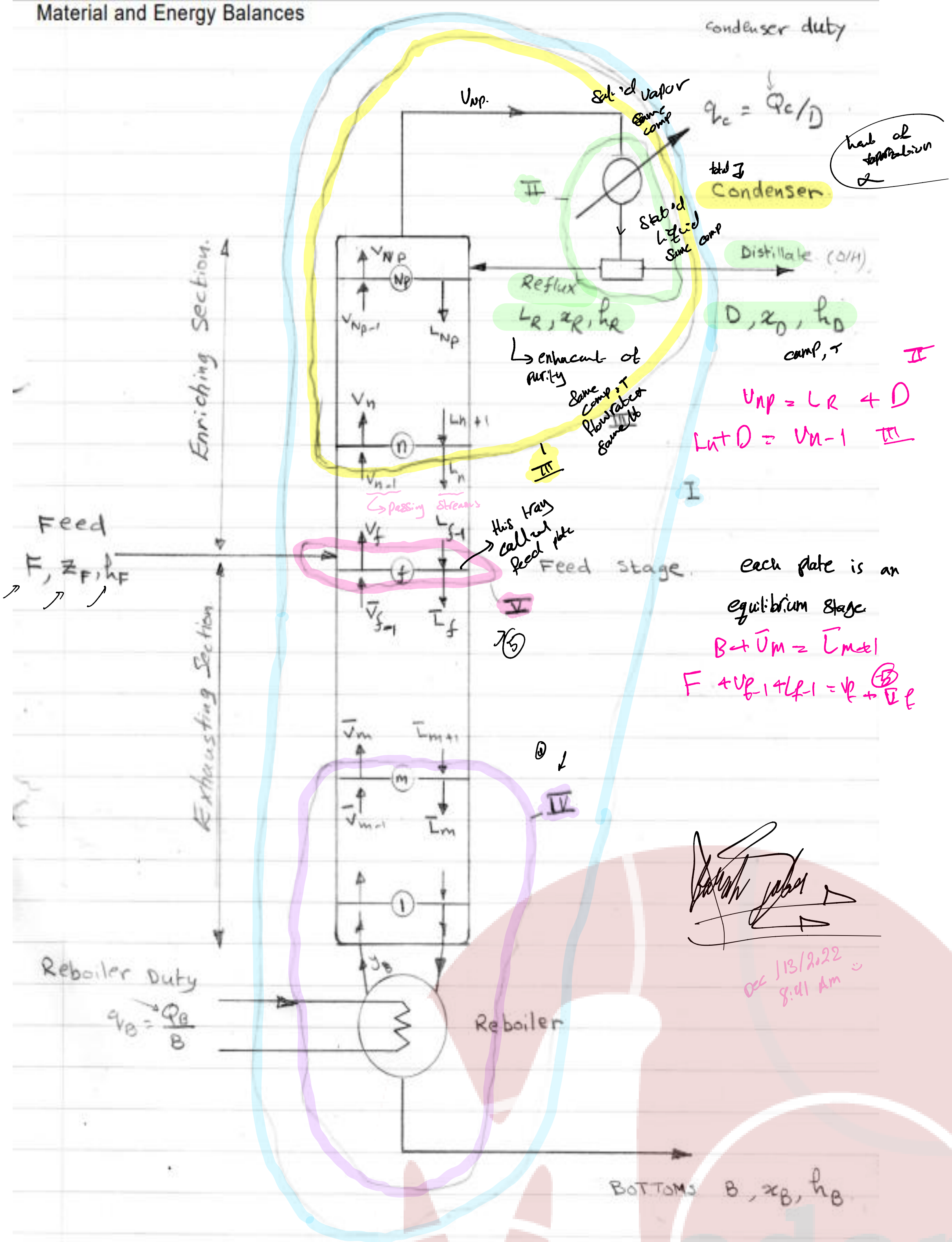
In some cases, multiple feeds and side streams are available. Some distillation columns consist of the top section only and they are called rectification columns.

Vapors rising from the top are totally (or partially) condensed in a condenser and some of the liquid is returned to the top of the column. This liquid is called reflux. The ratio of reflux to distillate is called reflux ratio or sometimes external ratio.

The liquid at the bottom is either heated by a coil placed at the base of the column or by an external reboiler; the more volatile material returns as vapor to the column. Inside the tower liquids and vapors are always at their bubble points and dew points respectively. Their flow rates are not constant along the column.





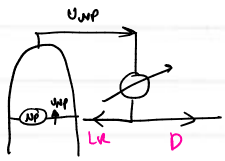

Partial condenser considered as a stage

Partial Reboiler considered a stage

Numbering of plates (trays) from Bottom (1) to top np

## Material Balances:

**Envelope I :** $F = D + B$




**Envelope II :** $V_{N_p} = L_R + D$

$R = \frac{L_R}{D}$

$DR = L_R$

$V_{np} = L_R + D$

$V_{np} = D(1+R)$

let $R = \frac{L_R}{D}$ = reflux ratio.

$$V_{N_p} = RD + D$$

$$V_{N_p} = D(R+1)$$ 2 b

(D → Distillate)

$D = V_{n-1} - L_n$

$B + \bar{V}_m = \bar{L}_{m+1}$

$B = \bar{L}_{m+1} - \bar{V}_m$

**Envelope III : (Down to stage n)**
↳ for top.

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$$V_{n-1} = L_n + D$$ 3 a

$$V_{n-1} - L_n = D$$ [Net Flow : Const] of vap up 3 b

↳ difference is Always constant.

**Envelope IV : (up to stage m).**
↳ for Bottom.

$$\bar{L}_{m+1} = \bar{V}_m + B$$ 4 a

(indication for Bottom)

$$\bar{L}_{m+1} - \bar{V}_m = B$$ [net flow : const] of liq

↳ Same as top. (constant)

**Envelope V : (Feed stage)**

in / out

$$F + L_{f+1} + \bar{V}_{f-1} = V_f + \bar{L}_f$$

## Component Balances: (For any component)

**Envelope I**

$$z_F \cdot F = x_D \cdot D + x_B \cdot B$$

→ composition of distillate

$$x_D = \frac{z_F \cdot F - x_B \cdot B}{F - B}$$

$$x_B = \frac{z_F \cdot F - x_D \cdot D}{F - D}$$

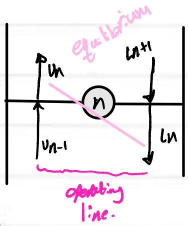
## Enriching Section (Envelope III)

$$y_{n-1} \cdot V_{n-1} = x_n \cdot L_n + D \cdot x_D \qquad [D x_D : \text{net Flow}]$$

$$y_{n-1} = \frac{L_n}{V_{n-1}} \cdot x_n + \frac{D x_D}{V_{n-1}}$$

↳ top.

op. line for enriching section




## Stripping Section (Envelope IV)

$$x_{m+1} \cdot \bar{L}_{m+1} = y_m \cdot \bar{V}_m + x_B \cdot B$$

$$y_m = \frac{\bar{L}_{m+1}}{\bar{V}_m} x_{m+1} - \frac{B}{\bar{V}_m} \cdot x_B$$

↳ Bottom.

## Enthalpy Balances (calculation of $Q_c$, $Q_B$)

→ heat duty per unit distillate ($Q_c$)

→ reboiler duty per unit waste or Bottom product ($Q_B$)

**Envelope II: (condenser duty)**

→ removing energy for condensing ($Q_c$).

$$V_{N_p} \cdot h_{y N_p} = Q_c + L_R \cdot h_R + D h_D$$

$$Q_c = V_{N_p} \cdot h_{y N_p} - D(R \cdot h_R + h_D)$$

→ equal because they have the same T, P, composition, it but different flow rate ($h_R$ and $h_D$)

$$Q_c = D\left[(R+1) \cdot h_{V,N_p} - (R h_R + h_D)\right]$$

$(R \uparrow Q_c \uparrow)$

**Rate of cooling Water:**

$$\dot{m} = \frac{Q_c}{C_p \Delta t}$$

→ enthalpy / mass.

$$\Delta t = T_{out} - T_{in}$$

Temp condenser

$$T_{in} = 20 \sim 25°C$$

كلما زاد درجة الحرارة بدي ماي داخل للتبريد أكثر

## Envelope I:

$$F \cdot h_F + q_B = D h_D + B h_B + Q_c + \cancel{Q_L}$$

negligible if well insulated [Total losses]

$$q_B = D\left(h_D + \frac{Q_c}{D}\right) + B \cdot h_B - F \cdot h_F + Q_L$$

first we have to calculate $Q_c$

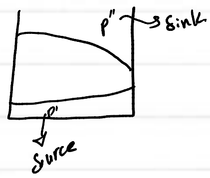
then $Q_B \Rightarrow$.

## Envelope III

$$V_{n-1} \cdot h_{V_{n-1}} - L_n \cdot h_{L_n} = D\left(\frac{Q_c}{D} + h_D\right) = \Delta_D \quad (p')$$

$$V_{n-1} \cdot h_{V_{n-1}} - L_n h_{L_n} = D(q_c + h_D)$$

= net flow.




## Envelope IV:

$$\bar{L}_{m+1} \cdot h_{\bar{L}, m+1} - \bar{V}_m \cdot h_{\bar{V}, m} = B\left(h_B - \frac{q_B}{B}\right) = \Delta_W \quad (p'')$$

$$L_{m+1} h_{m+1} - V_m \cdot h_{V_m} = B(h_B - q_B)$$

= net flow.

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The above relations together with equilibrium relations give the necessary equations for the design of distillation columns.

At the end of design calculations, one hopes to have

- $N_p$
- Product flow rates, compositions
- Operating conditions (T, P)
- $Q_c$, $Q_B$

**Specifications:**

Two types of problems may be encountered with any equipment, in this ca distillation column:

- Design Problem
- Simulation Problem → rating

Design Problem: Desired separation is set and column is designed that will achieve this separation

Simulation Problem: Column is already built and the procedure should predict how much separation can be achieved for a given feed.

For both problems we usually specify:

- Column pressure, (sets equilibrium data)
- Feed composition
- Feed flow rate.
- Feed temperature and pressure (or enthalpy or quality – state of the feed $L_f$ and $V_f$).  
  → feed quality: → liquid may be zero; → vapor may be zero; → mixture
- Reflux temperature or enthalpy (usually saturated liquid) (Simulation Problem).

Other variables which must be set for adequate specification.
For a binary system the most usual specifications and the resulting calculated variables are:

| Specifications | Calculated variables |
|---|---|
| Design:<br>- $x_D$, $x_B$ (mvc)<br>- R ($L_R$/D) | D, B, $Q_c$, $Q_B$, $N_p$, feed location.<br>Column Diameter. |
| Simulation:<br>- Np, feed location<br>- column diameter<br>- Reboiler size (gives max vapor)<br>and<br>- $x_D$ and $x_B$<br>or<br>- R, $x_D$ (or $x_B$)<br>or<br>- $x_D$ (or $x_B$), $V=V_{max}$ | <br><br><br><br>R, B, D, $Q_c$, $Q_R$ Check $v<v_{max}$<br><br>$x_B$ (or $x_D$),B,D,$Q_c$,$Q_R$ Check $v<v_{max}$<br><br>R, $x_B$ (or $x_D$), B, D, $Q_c$, $Q_R$ Check $v<v_{max}$ |

Two approaches may be used:

- **Approximate:** simple sequential, (adequate in many cases), graphs, simple algorithms
- **Rigorous:** requires detailed stage to stage calculations (simulation), handles all situations ⇒ useful in Hysys valid for multiple component

Not having Assumptions

## Binary Rectification:

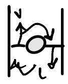
### Approximate Methods:

Simplifying assumptions:

1. Constant molal (→ number of moles) overflows: valid if (over range of temperature and Pressure):

   - Molar heats of vaporization of both species of the binary system are equal
   - Heats of mixing, stage heat losses and sensible heat changes of both liquid and vapour are negligible

   This assumption indicates that every mole of condensing vapour vaporizes exactly ONE mole of liquid.

   This means that on molal basis in each section of a column;

   $$\frac{Moles of\ Liquid}{Moles\ of\ Vapor} = constatnt\ ;\quad Ln = L\ ,\ \ Vn = V\ \ and\ \bar{L}_m = \bar{L}\ ,\ \bar{V}_m = \bar{V}$$

   This results in a straight line operating line for each section.

2. Distillation takes place at constant pressure (small variations of pressure inside the column)

stage to stage

### Plate to plate calculations: Lewis-Sorel Method

Consider the following case for the distillation of a binary mixture:

Given Information:

$F, \vec{Z}, x_D, x_B, R, P_{Column}$

Feed, Distillate, Bottoms product and reflux are saturated liquids at their bubble points. → feed is also stated if not we'll have a mixture

Required: Np and Feed location.

$F = D + B$

$z_f . F = x_D D + x_B B$

$L = RD$

$\bar{L} = L + F$

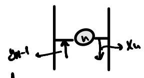
$\bar{V} = V$

$y_{n-1} = \frac{L}{V} x + \frac{x_D . D}{V}$

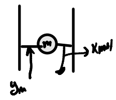
Analysis:

- Operating lines give relations between vapour and liquid compositions of passing streams:

  Rectifying Section: $$y_{n-1} = \frac{L}{V} . x_n + \frac{D}{V} . x_D$$ → net flow up


  Stripping Section: $$y_m = \frac{\bar{L}}{\bar{V}} . x_{m+1} - \frac{B}{\bar{V}} . x_B$$ → net flow down.


- Equilibrium data give relations between liquid and vapour compositions leaving a stage

$$x_n = f(y_n)$$

+ Material Balance ⇒

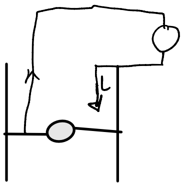
$$y_{n-1} = \frac{L}{V} x_n + x_D \frac{D}{V}$$

$$y_m = \frac{\bar{L}}{\bar{V}} x_{m+1} - \frac{B}{\bar{V}} x_B$$

4 eq.

## Calculation Procedure:

- Use column balances to calculate D, B, L and V, $\bar{L}$, $\bar{V}$



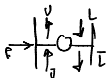
$D = \frac{F(z_F - x_B)}{(x_D - x_B)}$ ⇒ Material Balance ⇒

$B = F - D$ or $B = D\left(\frac{z_F - x_D}{x_B - z_F}\right)$

$L = L_R = RD$

$V = L_R(L) + D \Rightarrow D(R+1)$

feed plate:

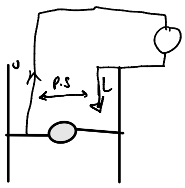


$\bar{L} = F + L$ ↘ because the feed is All liquid (Sat'd L)

$\bar{V} = V$

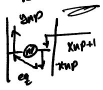
- Starting with top specification, find ..$y_{np}$..... from ......operating..line.........

$y_{np} = \frac{L}{V} x_{np+1} + \frac{D}{V} \cdot x_D$



$\boxed{y_{np} = x_{np+1}}$ $n = np+1$

$x_{np+1} = x_D = y_{np}$ ⇒ total condenser.



- Find..$x_{np}$....from ..equilibrium..eq.........

$\boxed{y_{np} = K \cdot x_{np}}$ if they have same structure ↳ Law or equilibrium curve graph

- Repeat for stage ..np-1...

operating line $n = np$

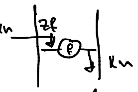
$y_{np-1} = \frac{L}{V} x_{np} + \frac{D}{V} \cdot x_D$

$y_{np-1}$ ⇒

then $y_{np-1} = K x_{np-1}$ ↳ find $(x_{np-1})$ ⇒

- Repeat for stages below until .$x_n$ calculated calculated is nearly the same as $z_f$...... ↳ from equilibrium

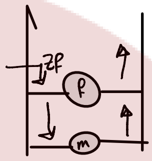
i.e $x_f = z_f = x_n$



for sat'd liquid feed $(x_f < z_f)$ Less more volatile component

Bottom section ↘

- Change ...operating..line..........., and starting with ..$m = f-1$..feed... ↙ stage below feed



operating line ⇒ calculate $(y_{f-1})$

$y_{f-1} = \frac{\bar{L}}{\bar{V}} x_f - \frac{B}{\bar{V}} \cdot x_B$

- Find..$x_{f-1}$....from ....eq..Data.......

equilibrium Data $(x_{f-1})$

$y_{f-1} = f(x_{f-1})$

- Repeat for stages below until .$x_m$..... calculated is nearly the same as ..$x_B$.....

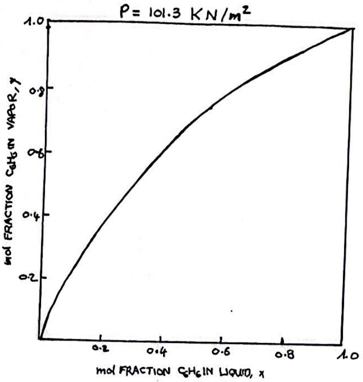
$x_m = f(y_m)$

or $x_w$

- Stop

Ex : Calculation of number of theoretical plates :- Lewis-Sorel method

Eqm. data :-




Operating conditions:-

Feed : binary mixture of benzene (40 mol%) + Toluene, at boiling poi[nt] ↳ sat'd Liquid

$x_D$.

Distillate : 90% benzene at bubble point (liqui[d]) from total condenser

↳ أكثر من هيك يعني ما نكمل

Bottom product : not more than 10% benzene

Reflux Ratio : R = 3

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**Solution :**

- Basis 100 kg-mole of feed
- Determine D and W from MB. → Material Balance

W = 62.5 Kmol ; D = 37.5 Kmol

- $L_n$ = 3 x 37.5 = 112.5 kmol ; G = 150 k-mol (D + $L_n$) ↳ vapor (يلي طالع من فوق) ; $L_m$ = 100 + 112.5 = 212.5 kmol (feed + $L_n$)

- Operating lines

rectifying section : $y_{n+1} = 0.75\,x_n + 0.225$

stripping section : $y_{m+1} = 1.415\,x_m - 0.042$

Equilibrium compositions on theoretical stages

| | RECTIFICATION | | STRIPPING | |
|---|---|---|---|---|
| | $y_n \equiv y_{np}$ | $x_n \equiv x_{np-1}$ | $y_m$ | $x_m$ |
| 1 | 0.9 = $x_D$. | 0.79 from op. | | |
| 2 | 0.818 | from equilibrium 0.644 | | |
| 3 | 0.708 | 0.492 | | |
| 4 | 0.594 | 0.382 → stop | | كيف بنلاقي هو؟ |
| 5 | | $x_n < x_F$ we have four stages | 0.498 | 0.298 |
| 6 | | at top section | 0.379 | 0.208 |
| 7 | | | 0.252 | 0.120 |
| 8 | | | 0.127 | 0.048 * ↳ below 10%. |
| 9 | | | | |

* Reboiler stage ∴ no. of plates = 7

$\alpha_{avg} = 2.75$