Distillation

Processos de Separação

LEQB

2023/2024

Summary

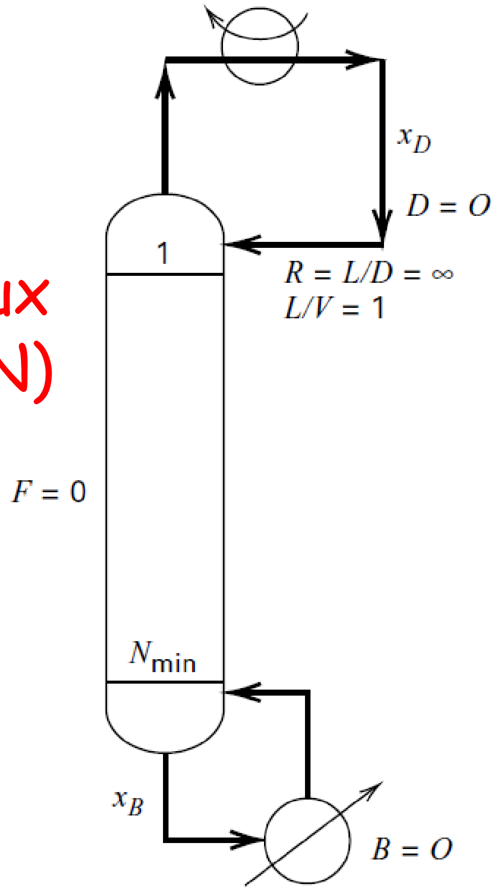
- Optimal reflux ratio
- Condenser and Boiler in a distillation column
- Energy Balances to a distillation column
- Stage efficiency
- Energetic balance to the feed stage

- **Optimal reflux ratio**

Reflux ratio, *R*

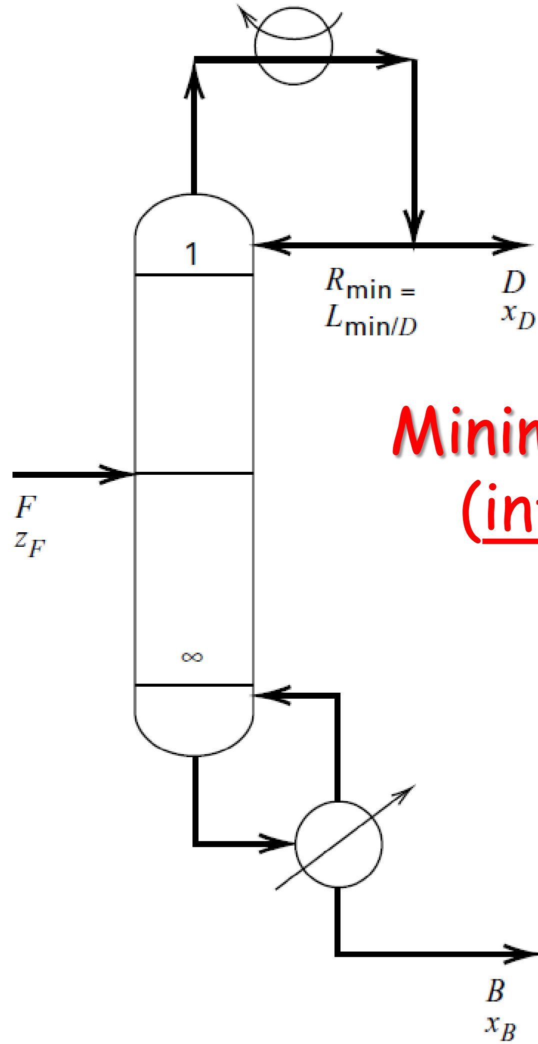
R= L/D

Total reflux
(<u>minimum</u> N)



↖ R ↘

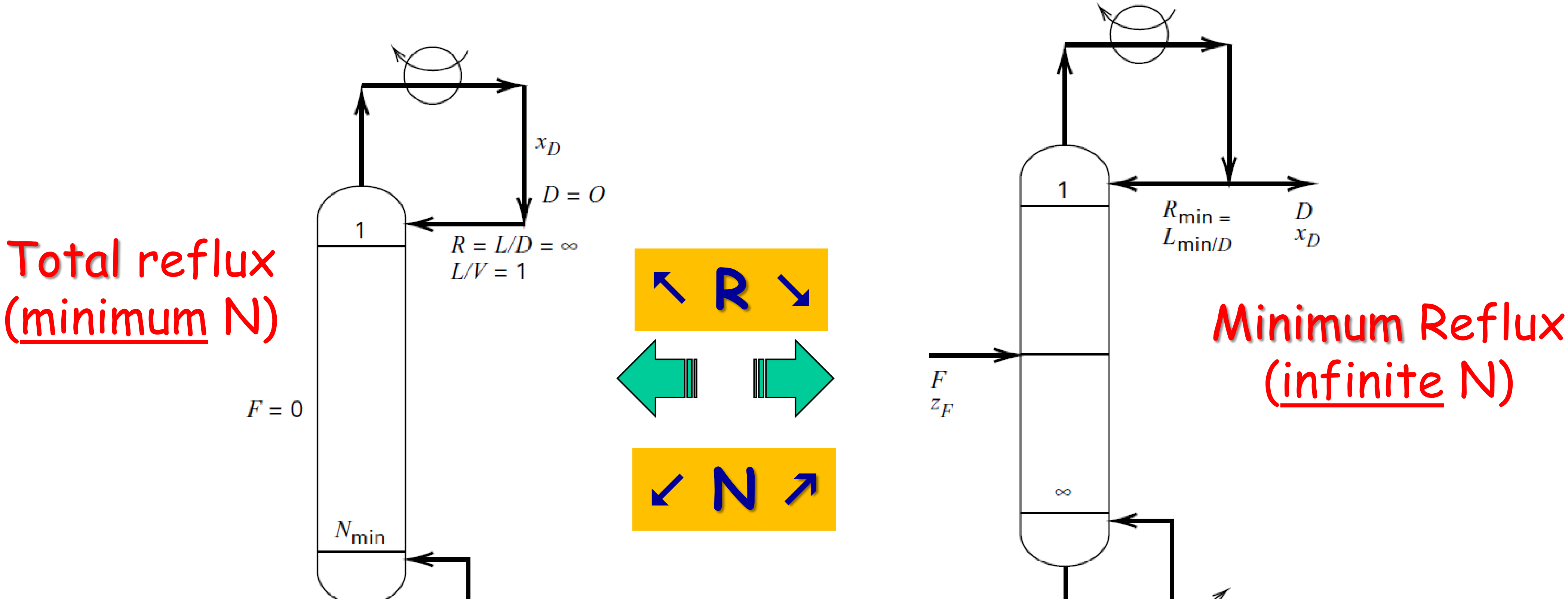
↙ N ↗



Minimum Reflux
(<u>infinite</u> N)

Reflux ratio, *R*

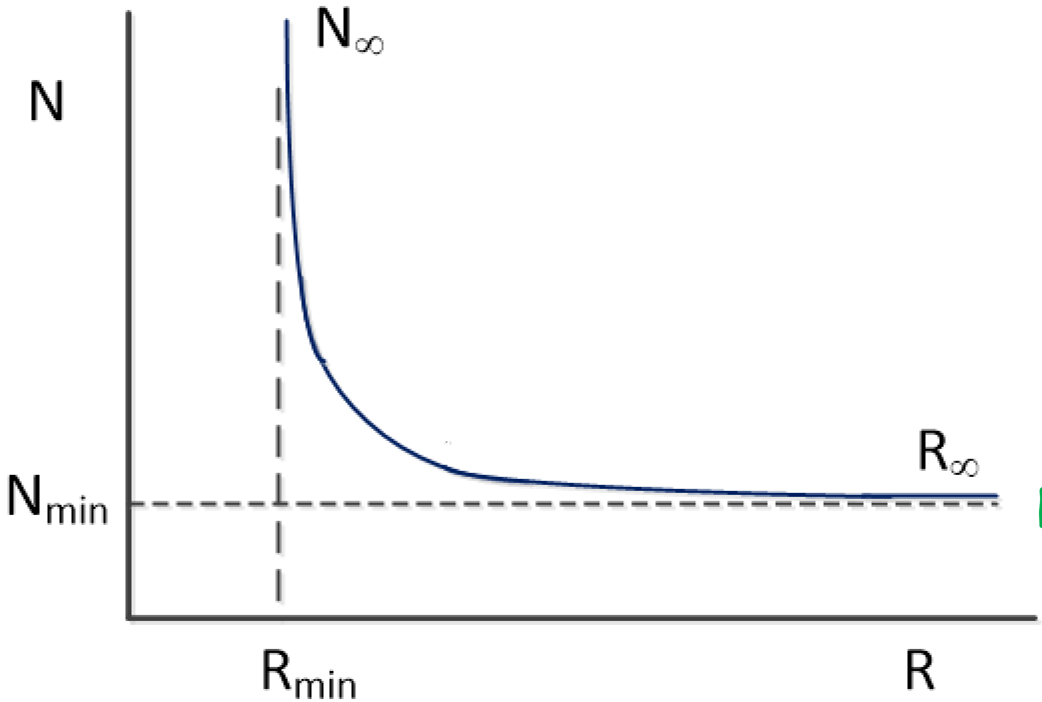
R= L/D



which value should we choose for the reflux ratio of a given distillation column, and how do we estimate that value ??

Reflux Ratio, R ↔ Number of stages, N

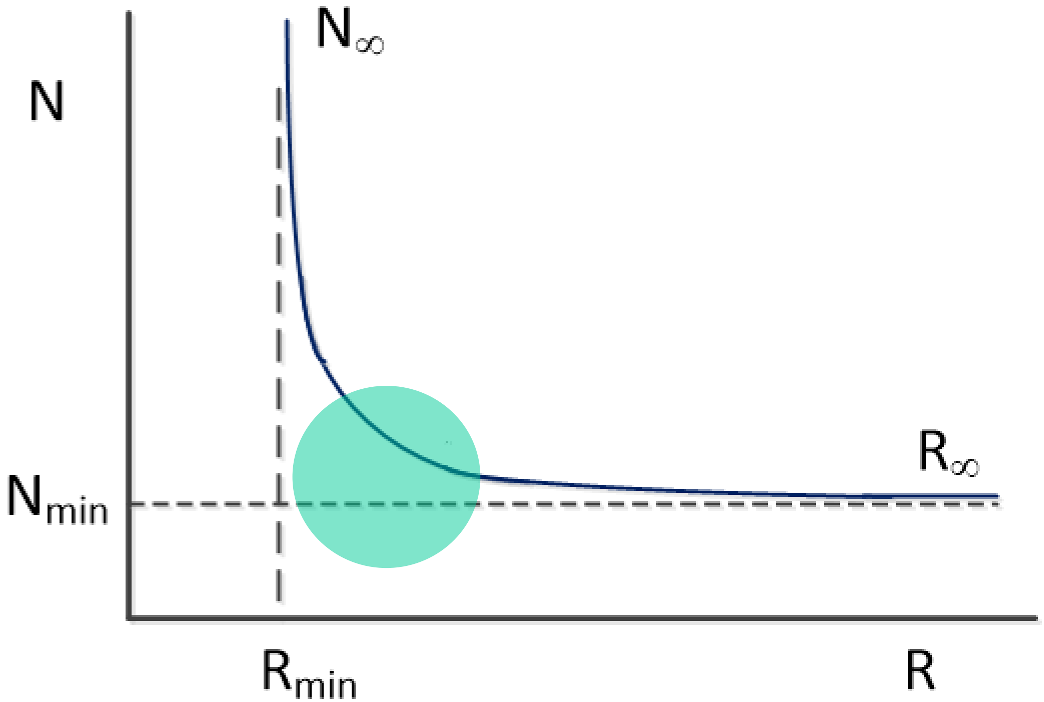
R minimum ↔ N infinite



R total ↔ N minimum

Reflux Ratio, R ↔ Number of stages, N

Optimum Reflux → $1.05 \text{ a } 1.5 \times R_{min}$



How?

- Material Balance
- Energy Balance
- Economic Analysis

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗ R/R_{min}	↘ Actual N
1.00	Infinite
1.05	29
1.14	21
1.23	18
1.32	16
1.49	14
1.75	13

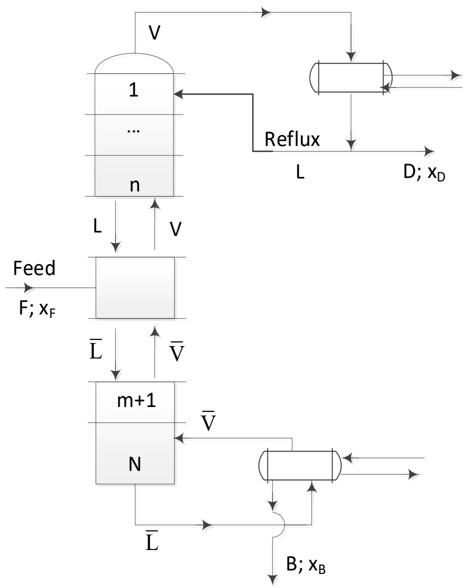
(Adapted from an example by Peters and Timmerhaus [6].)

Values of F, x_F, x_D, x_B and i are constants

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

$R \nearrow \Leftrightarrow L \nearrow$

System benzene + toluene



M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗ R/R_{min}	↘ Actual N	↗ Diam., ft	
1.00	Infinite	6.7	(204 cm)
1.05	29	6.8	
1.14	21	7.0	
1.23	18	7.1	
1.32	16	7.3	
1.49	14	7.7	
1.75	13	8.0	(243 cm)

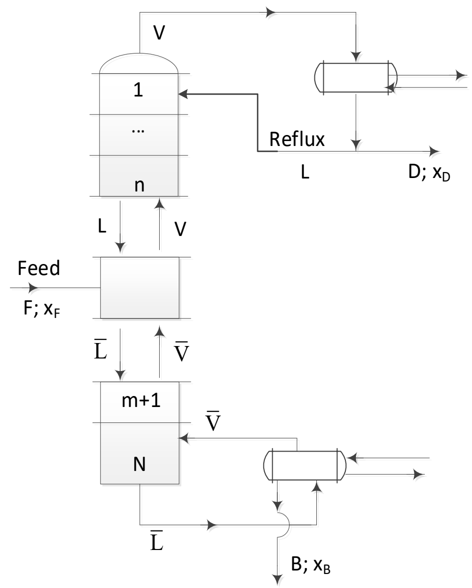
(Adapted from an example by Peters and Timmerhaus [6].)

System benzene + toluene

Values of F, x_F, x_D, x_B and *i* are constants

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

R ↗ ⇔ L ↗



M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗	↘	↗	↗	↗
R/R_{min}	Actual N	Diam., ft	Reboiler Duty, Btu/h	Condenser Duty, Btu/h
1.00	Infinite	6.7	9,510,160	9,416,000
1.05	29	6.8	9,776,800	9,680,000
1.14	21	7.0	10,221,200	10,120,000
1.23	18	7.1	10,665,600	10,560,000
1.32	16	7.3	11,110,000	11,000,000
1.49	14	7.7	11,998,800	11,880,000
1.75	13	8.0	13,332,000	13,200,000

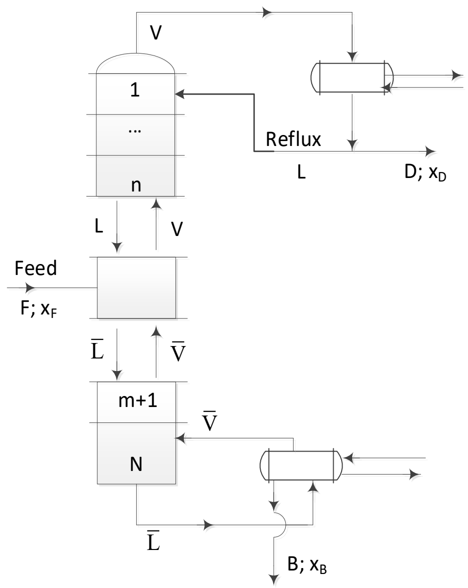
(Adapted from an example by Peters and Timmerhaus [6].)

Values of F, x_F, x_D, x_B and *i* are constants

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

R ↗ ⇔ L ↗

System benzene + toluene



M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗	↘	↗	↗	↗	Annualized Cost, \$/yr
R/R_{min}	Actual N	Diam., ft	Reboiler Duty, Btu/h	Condenser Duty, Btu/h	↘ Equipment
1.00	Infinite	6.7	9,510,160	9,416,000	Infinite
1.05	29	6.8	9,776,800	9,680,000	44,640
1.14	21	7.0	10,221,200	10,120,000	38,100
1.23	18	7.1	10,665,600	10,560,000	36,480
1.32	16	7.3	11,110,000	11,000,000	35,640
1.49	14	7.7	11,998,800	11,880,000	35,940
1.75	13	8.0	13,332,000	13,200,000	36,870

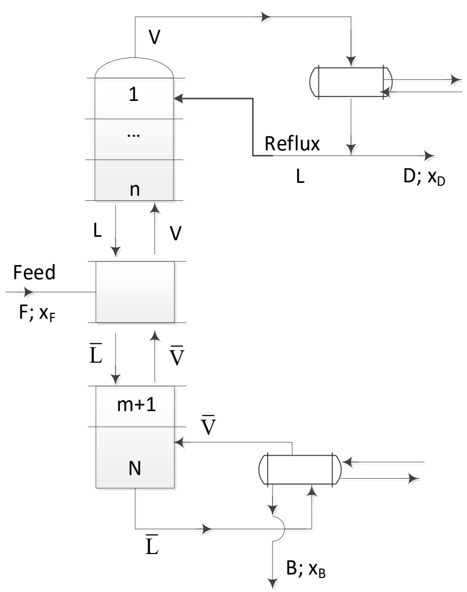
(Adapted from an example by Peters and Timmerhaus [6].)

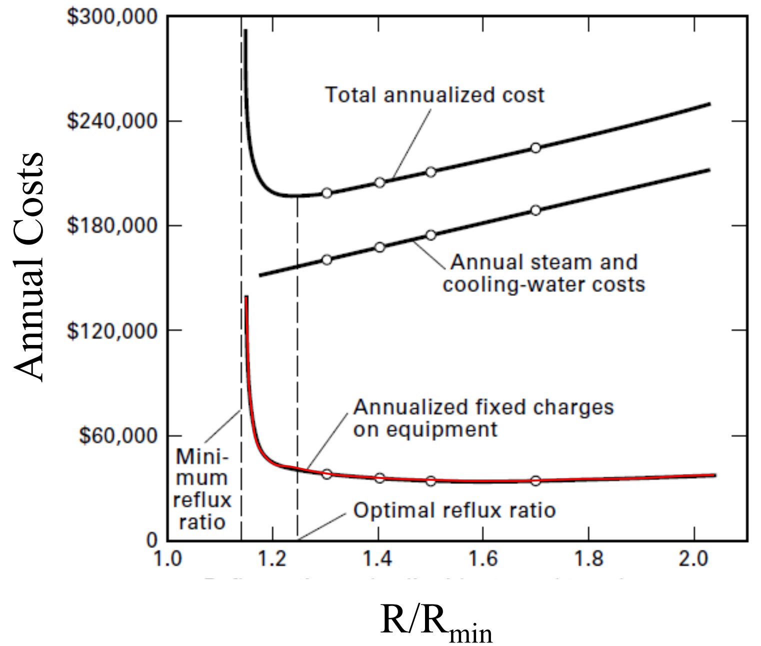
Values of F, x_F, x_D, x_B and *i* are constants

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

R ↗ ⇔ L ↗

System benzene + toluene





M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗	↘	↗	↗	↗	Annualized Cost, \$/yr		
					↘		↗
R/R_{min}	Actual N	Diam., ft	Reboiler Duty, Btu/h	Condenser Duty, Btu/h	Equipment	Cooling Water	Steam
1.00	Infinite	6.7	9,510,160	9,416,000	Infinite	17,340	132,900
1.05	29	6.8	9,776,800	9,680,000	44,640	17,820	136,500
1.14	21	7.0	10,221,200	10,120,000	38,100	18,600	142,500
1.23	18	7.1	10,665,600	10,560,000	36,480	19,410	148,800
1.32	16	7.3	11,110,000	11,000,000	35,640	20,220	155,100
1.49	14	7.7	11,998,800	11,880,000	35,940	21,870	167,100
1.75	13	8.0	13,332,000	13,200,000	36,870	24,300	185,400

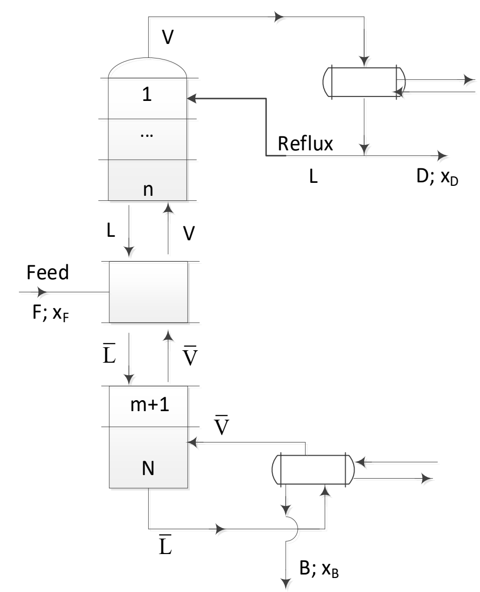
(Adapted from an example by Peters and Timmerhaus [6].)

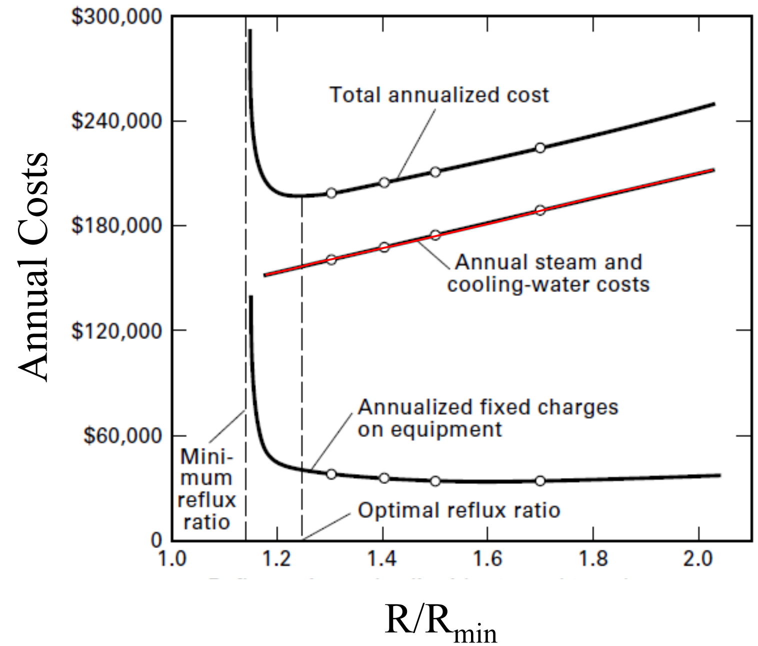
Values of F, x_F, x_D, x_B and i are constants

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

R ↗ ⇔ L ↗

System benzene + toluene





M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗ R/R_{min}	↘ Actual N	↗ Diam., ft	↗ Reboiler Duty, Btu/h	↗ Condenser Duty, Btu/h	Annualized Cost, \$/yr ↘ Equipment	Cooling Water	↗ Steam	Total Annualized Cost, \$/yr
1.00	Infinite	6.7	9,510,160	9,416,000	Infinite	17,340	132,900	Infinite
1.05	29	6.8	9,776,800	9,680,000	44,640	17,820	136,500	198,960
1.14	21	7.0	10,221,200	10,120,000	38,100	18,600	142,500	199,200
1.23	18	7.1	10,665,600	10,560,000	36,480	19,410	148,800	204,690
1.32	16	7.3	11,110,000	11,000,000	35,640	20,220	155,100	210,960
1.49	14	7.7	11,998,800	11,880,000	35,940	21,870	167,100	224,910
1.75	13	8.0	13,332,000	13,200,000	36,870	24,300	185,400	246,570

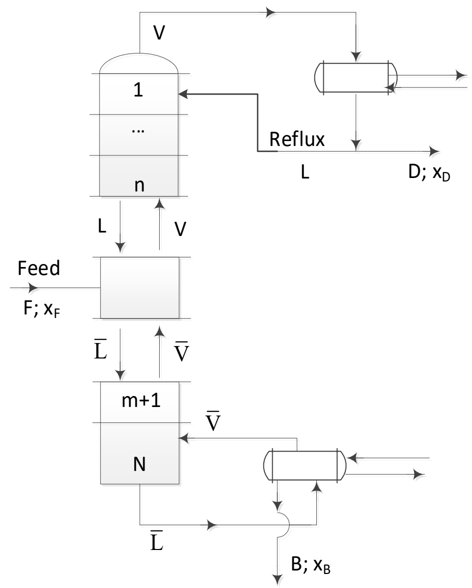
(Adapted from an example by Peters and Timmerhaus [6].)

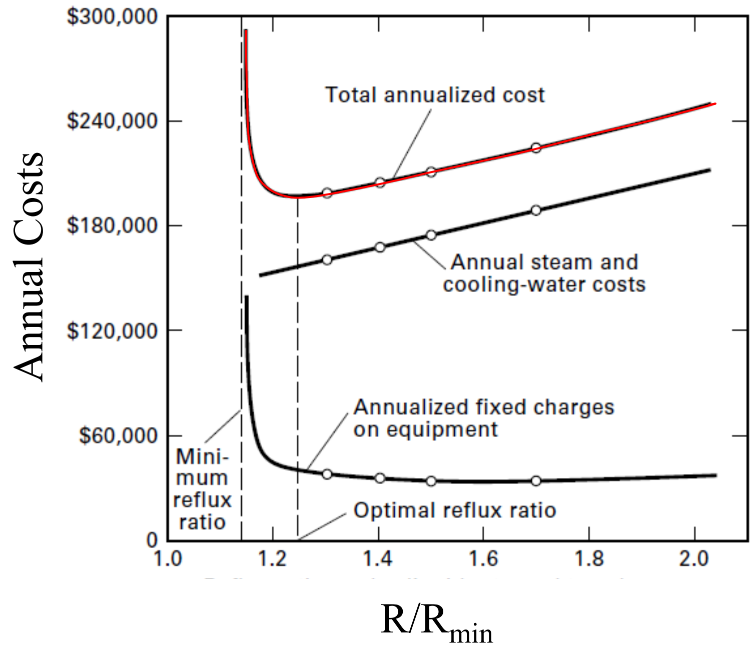
Values of F, x_F, x_D, x_B and i are constants

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

R ↗ ⇔ L ↗

System benzene + toluene





M Peters, K Timmerhaus
PLANT DESIGN AND ECONOMICS FOR CHEMICAL ENGINEERS, 4th ed, 1991

Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗	↘	↗	↗	↗	Annualized Cost, $/yr			
R/R_{min}	Actual N	Diam., ft	Reboiler Duty, Btu/h	Condenser Duty, Btu/h	↘ Equipment	Cooling Water ↗	Steam	Total Annualized Cost, $/yr
1.00	Infinite	6.7	9,510,160	9,416,000	Infinite	17,340	132,900	Infinite
1.05	29	6.8	9,776,800	9,680,000	44,640	17,820	136,500	198,960
1.14	21	7.0	10,221,200	10,120,000	38,100	18,600	142,500	199,200
1.23	18	7.1	10,665,600	10,560,000	36,480	19,410	148,800	204,690
1.32	16	7.3	11,110,000	11,000,000	35,640	20,220	155,100	210,960
1.49	14	7.7	11,998,800	11,880,000	35,940	21,870	167,100	224,910
1.75	13	8.0	13,332,000	13,200,000	36,870	24,300	185,400	246,570

(Adapted from an example by Peters and Timmerhaus [6].)

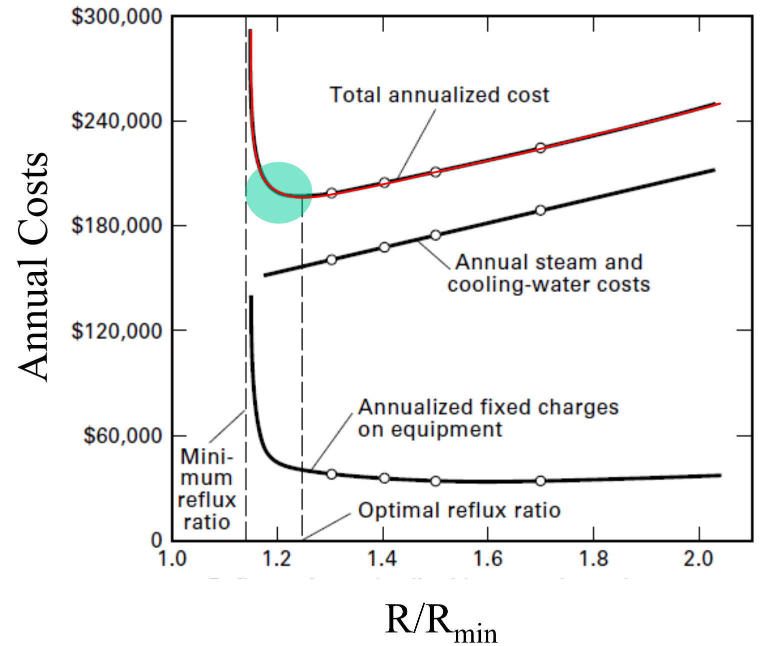
Values of F, x_F, x_D, x_B and *i* are constants

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

R ↗ ⇔ L ↗

System benzene + toluene

$$\Rightarrow \left.\frac{R}{R_{min}}\right]_{optimum} \cong 1.1$$



Reflux Ratio, R ↔ Number of stages, N

Effect of Reflux Ratio on Annualized Cost of a Distillation Operation

↗ R/R_{min}	↘ Actual N	↗ Diam., ft	↗ Reboiler Duty, Btu/h	↗ Condenser Duty, Btu/h	Annualized Cost, $/yr ↘ Equipment	Cooling Water ↗	Steam	Total Annualized Cost, $/yr
1.00	Infinite	6.7	9,510,160	9,416,000	Infinite	17,340	132,900	Infinite
1.05	29	6.8	9,776,800	9,680,000	44,640	17,820	136,500	198,960
1.14	21	7.0	10,221,200	10,120,000	38,100	18,600	142,500	199,200
1.23	18	7.1	10,665,600	10,560,000	36,480	19,410	148,800	204,690
1.32	16	7.3	11,110,000	11,000,000	35,640	20,220	155,100	210,960
1.49	14	7.7	11,998,800	11,880,000	35,940	21,870	167,100	224,910
1.75	13	8.0	13,332,000	13,200,000	36,870	24,300	185,400	246,570

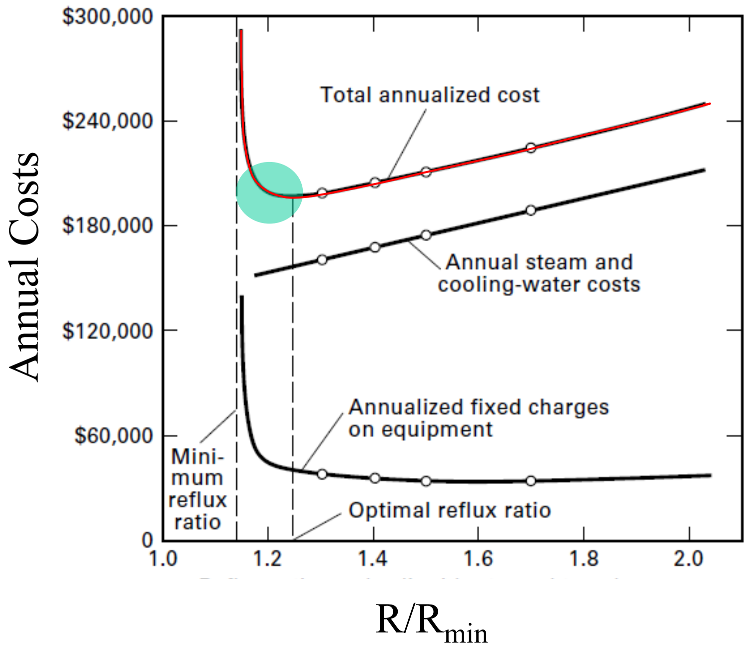
(Adapted from an example by Peters and Timmerhaus [6].)

System benzene + toluene

Values of F, x_F, x_D, x_B and i are constants

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

R ↗ ⇔ L ↗



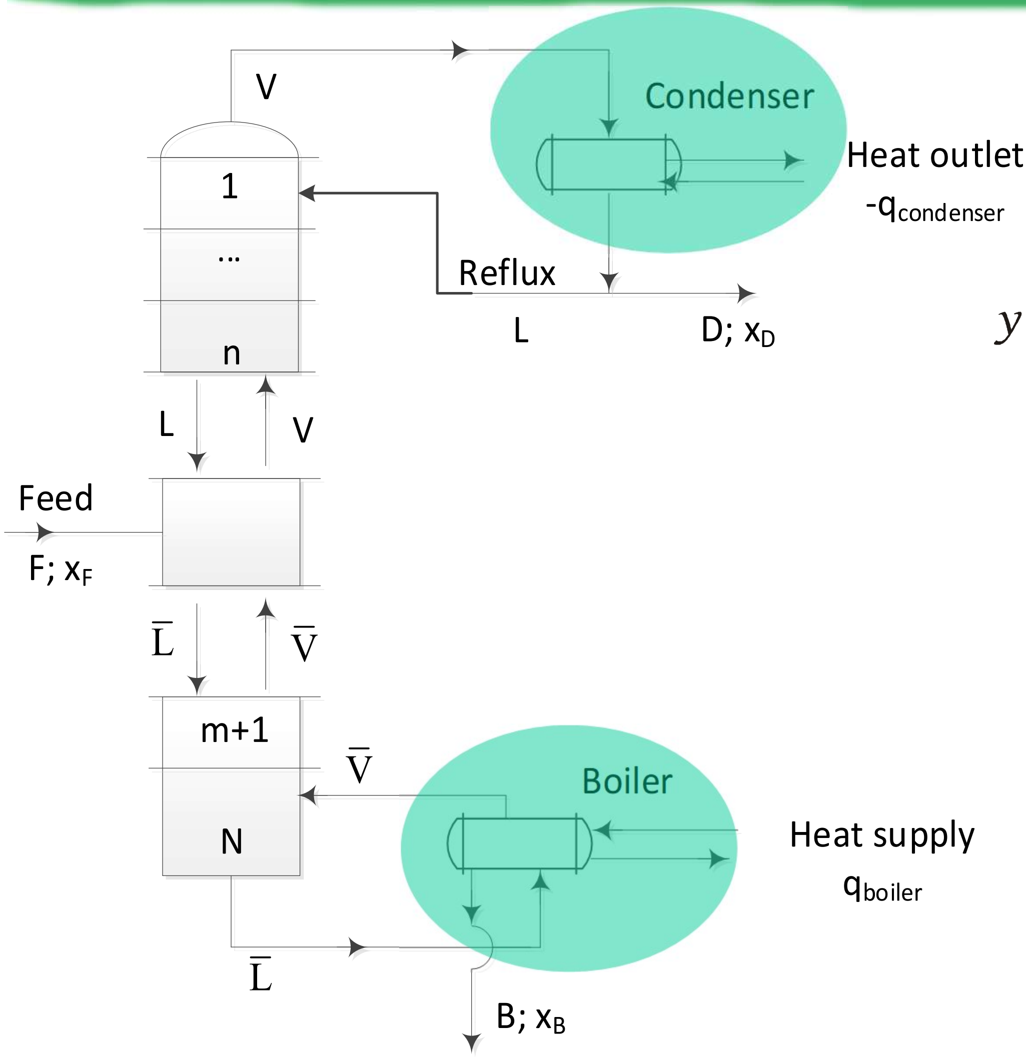
$$\left.\frac{R}{R_{min}}\right]_{optimum} \cong 1.1$$

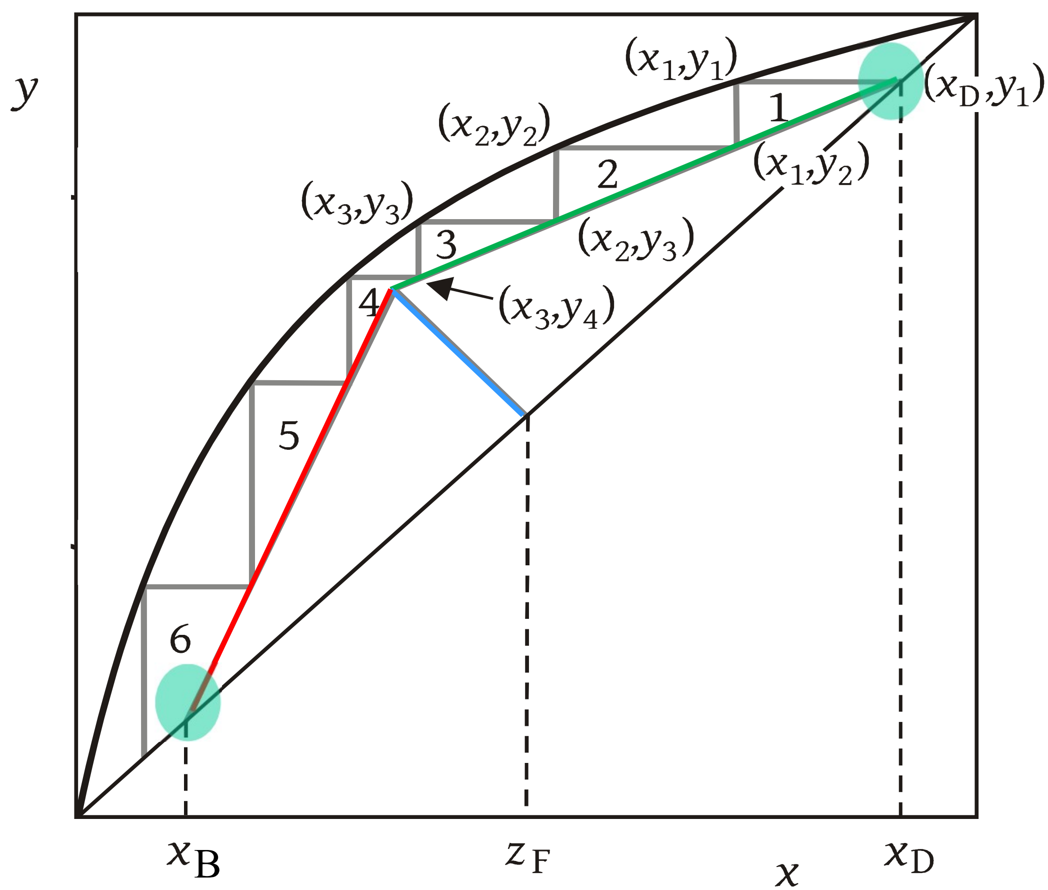
In this example!!!

Summary

- Influence of the reflux ratio on the efficiency of a distillation column
- Optimal reflux ratio
- Condenser and Boiler in a distillation column
- Energy Balances to a distillation column
- Energetic balance to the feed stage
- Stage efficiency

Distillation – McCabe – Thiele Method





Condenser and boiler

A distillation unit includes among others:

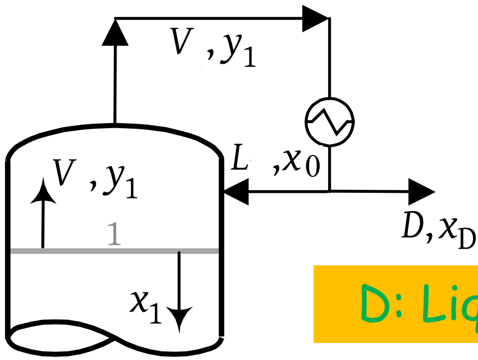
1) The distillation **column**
2) The **condenser**
3) The **boiler**

The McCabe-Thiele method calculates the **total** number of <u>equilibrium</u> stages of the distillation **unit**

Depending on the conditions, this number may include or not the boiler and the condenser

Condenser

Total Condenser

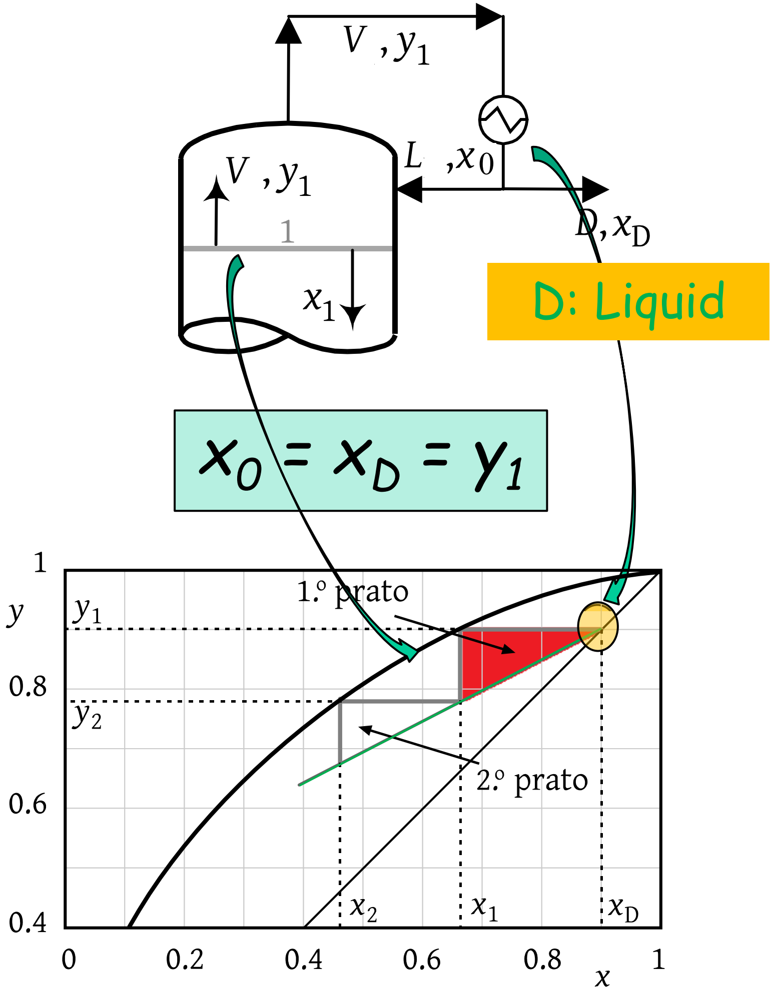


D: Liquid

$$x_0 = x_D = y_1$$

Condenser

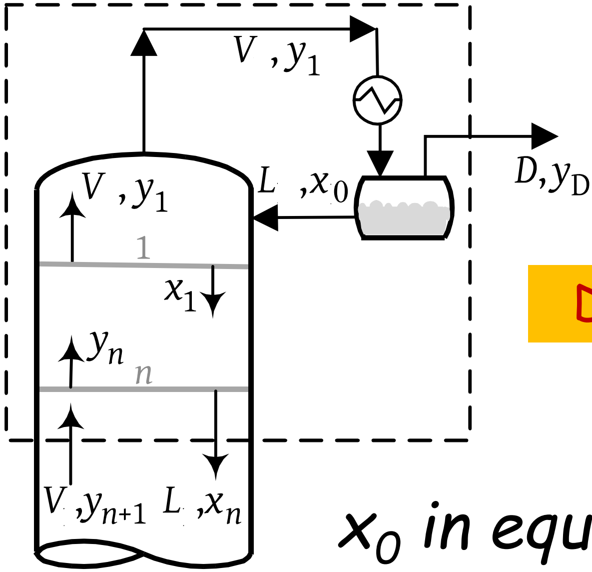
Total Condenser



The Condenser is not a equilibrium stage!

Condenser

Partial Condenser

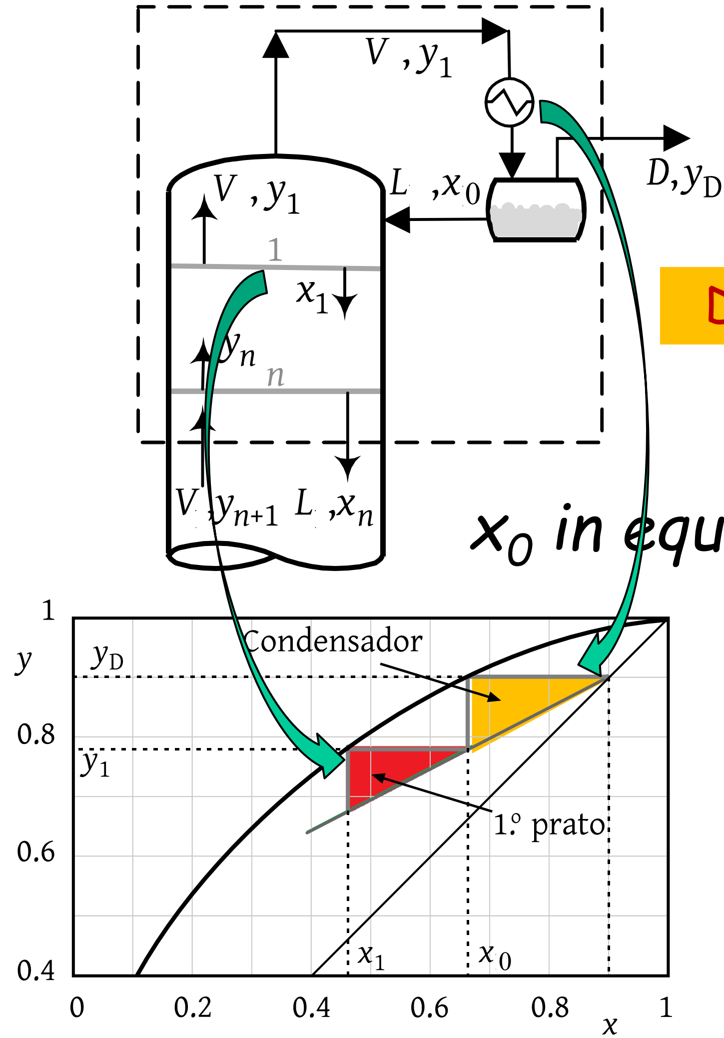


D: Vapor

x_0 in equilibrium with y_D

Condenser

Partial Condenser



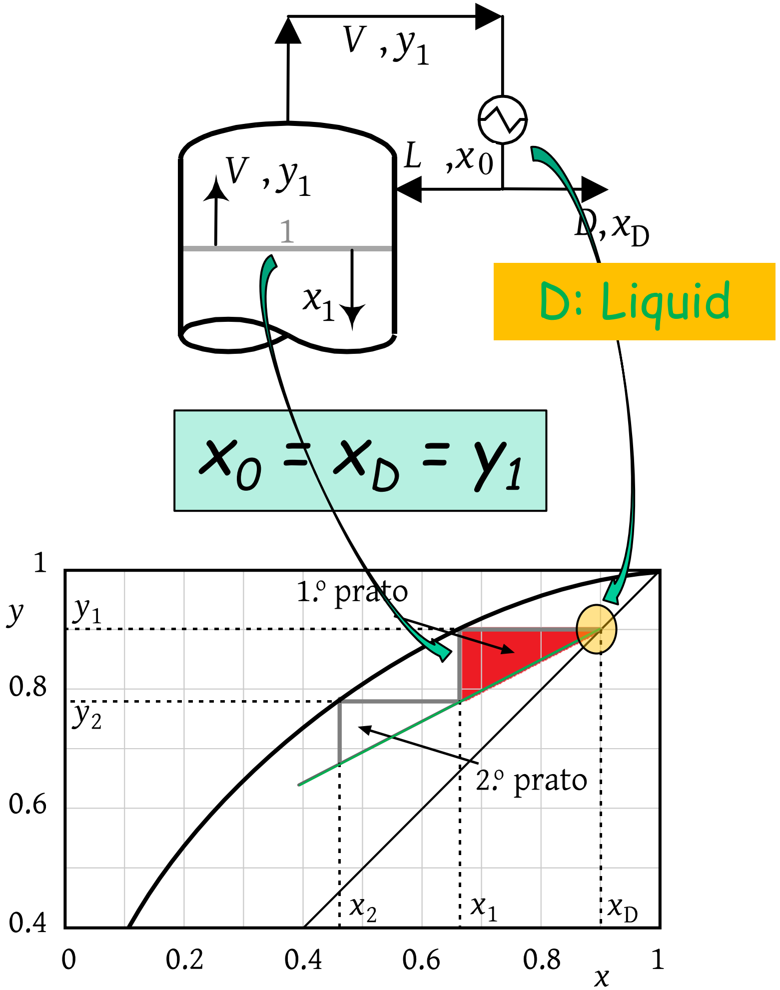
D: Vapor

x_0 in equilibrium with y_D

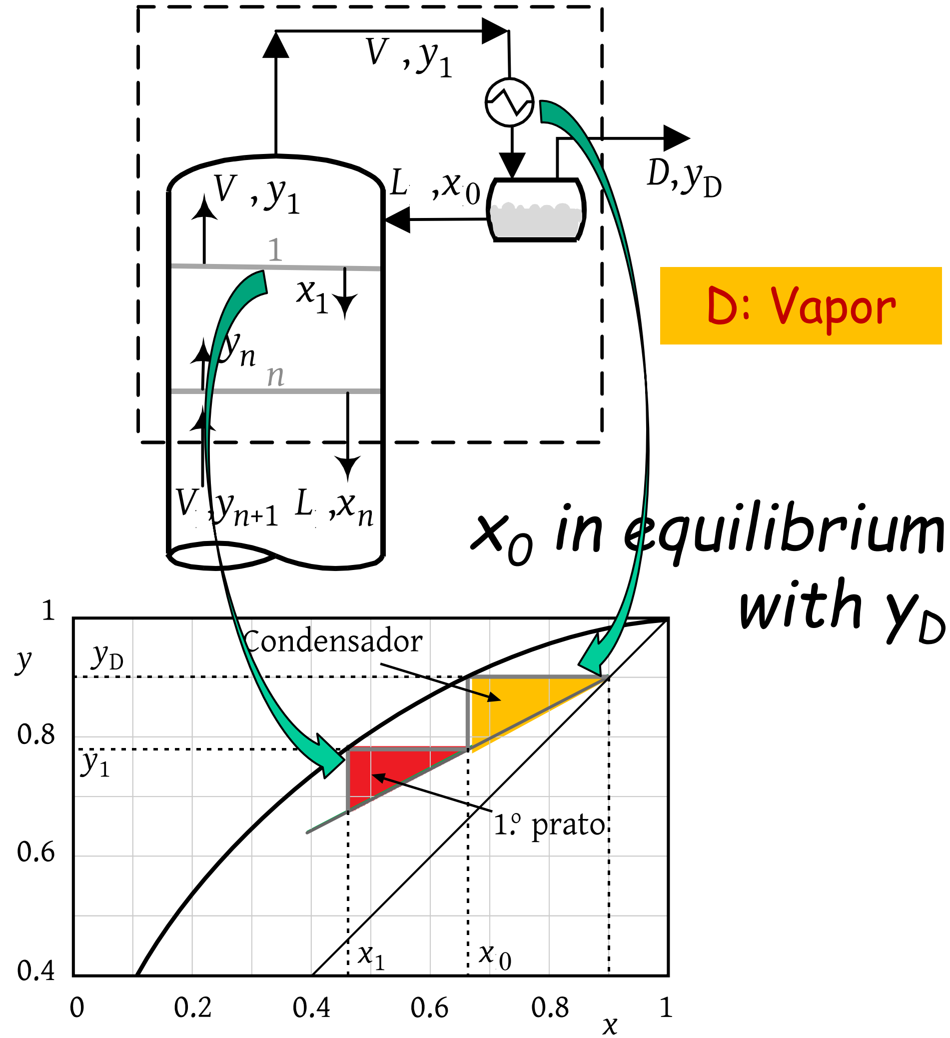
The Condenser is an equilibrium stage!

Condenser

Total Condenser



Partial Condenser



Condenser is not an equilibrium stage

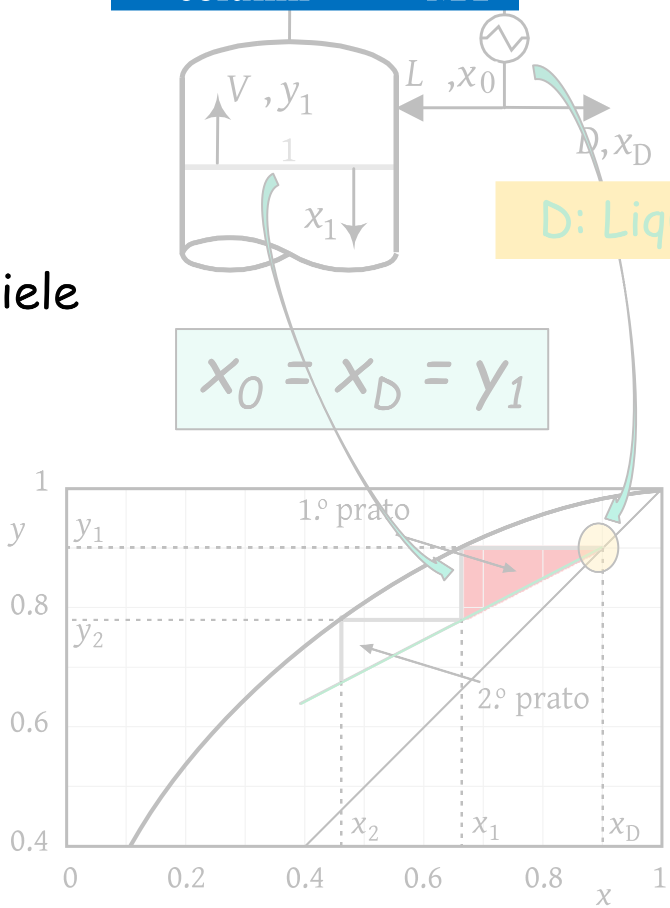
Condenser is an equilibrium stage

Condenser

Total Condenser

$N_{column} = N_{MT}$

MT: McCabe-Thiele



D: Liquid

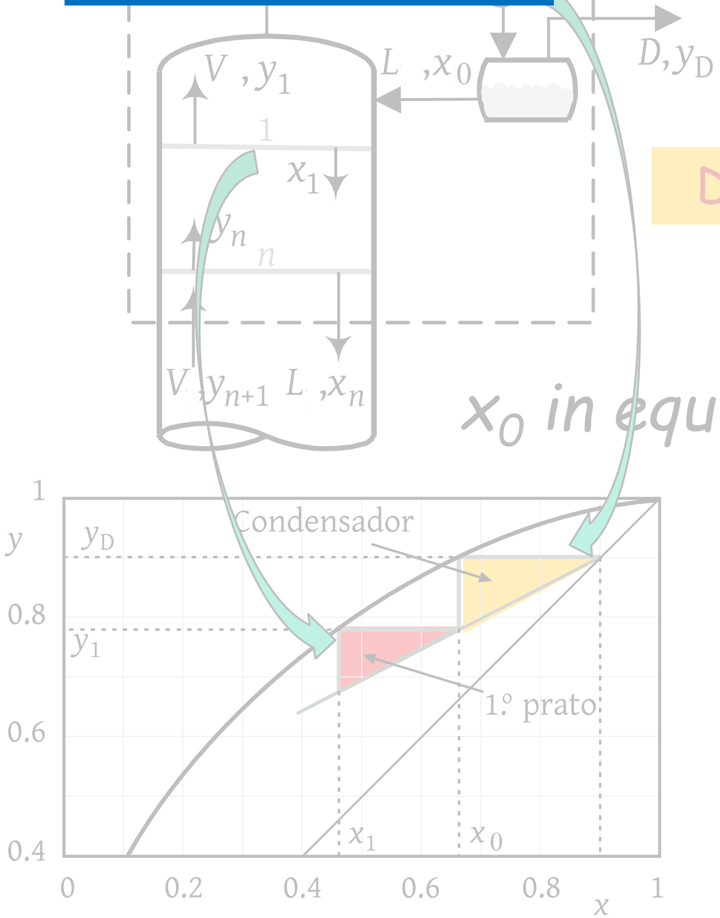
$x_0 = x_D = y_1$

Condenser is not an equilibrium stage

Partial Condenser

$N_{column} = N_{MT} - 1$

Condenser



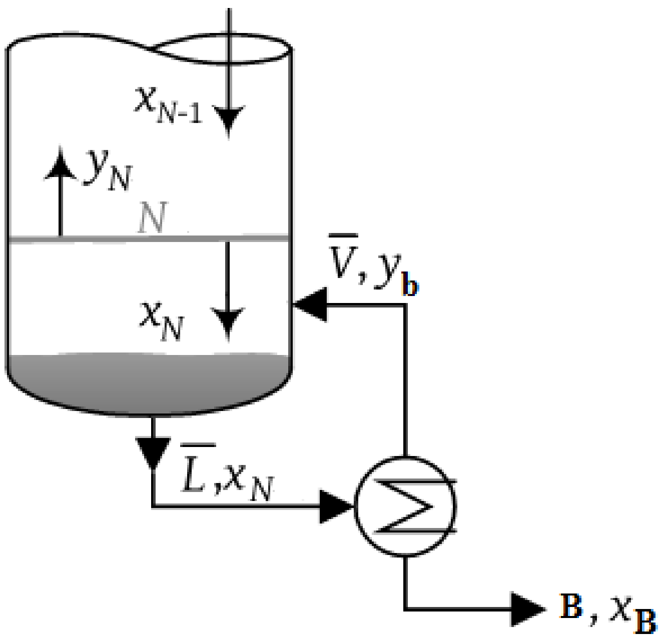
D: Vapor

x_0 in equilibrium with y_D

Condenser is an equilibrium stage

Boiler

Partial Boiler

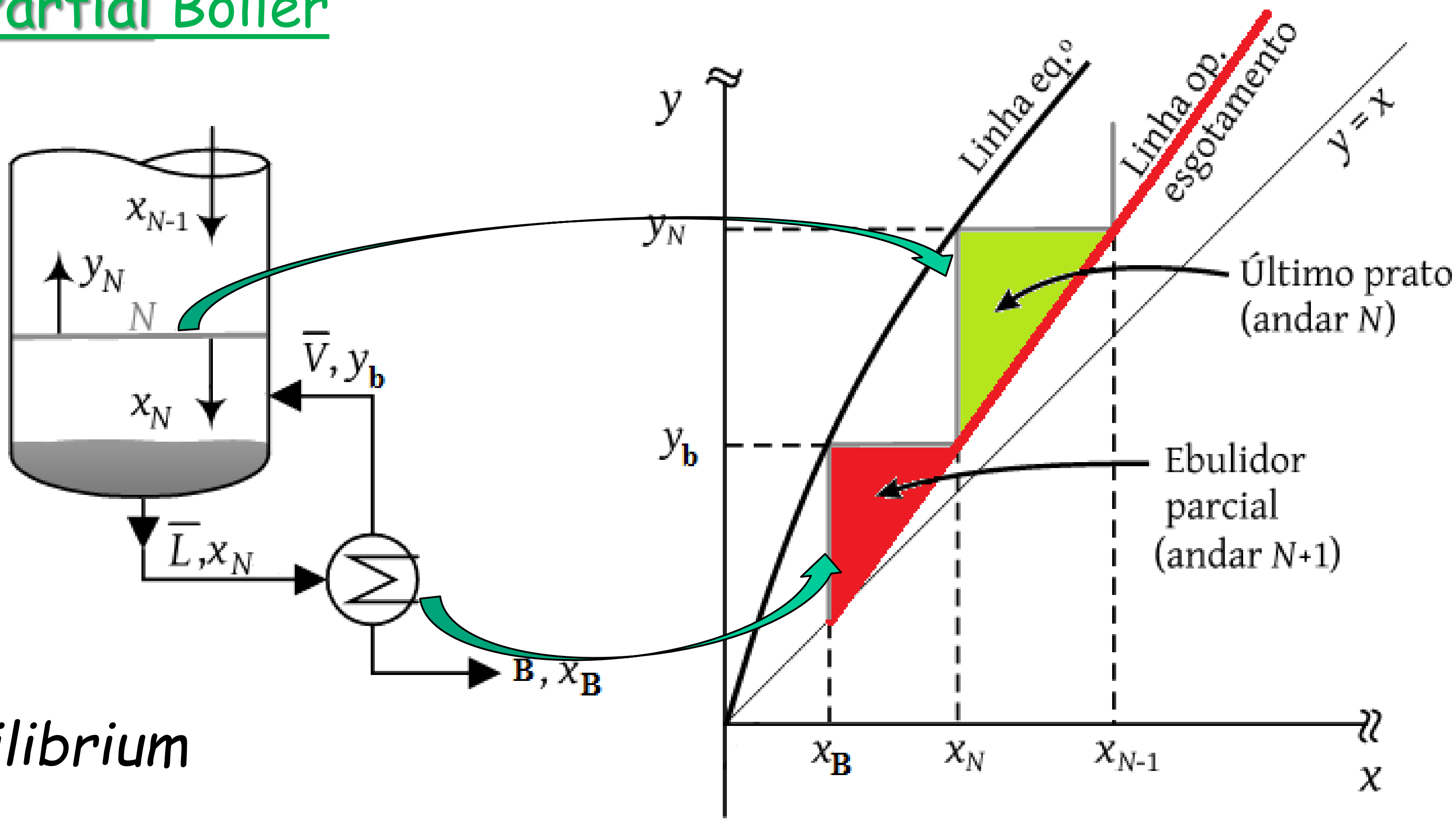


B: Liquid

x_B in equilibrium with y_b

Boiler

Partial Boiler



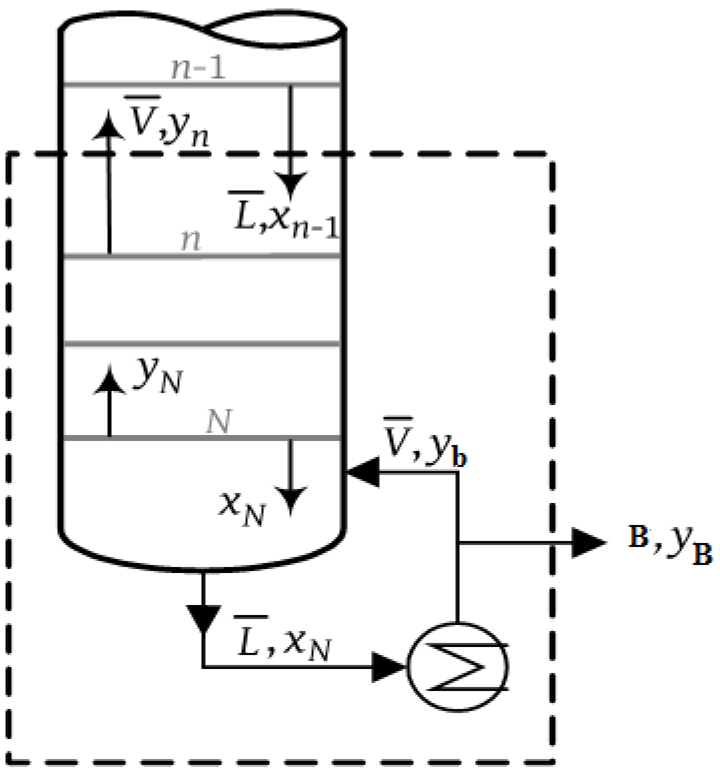
B: Liquid

x_B in equilibrium with y_b

Boiler is an equilibrium stage

Boiler

Total Boiler

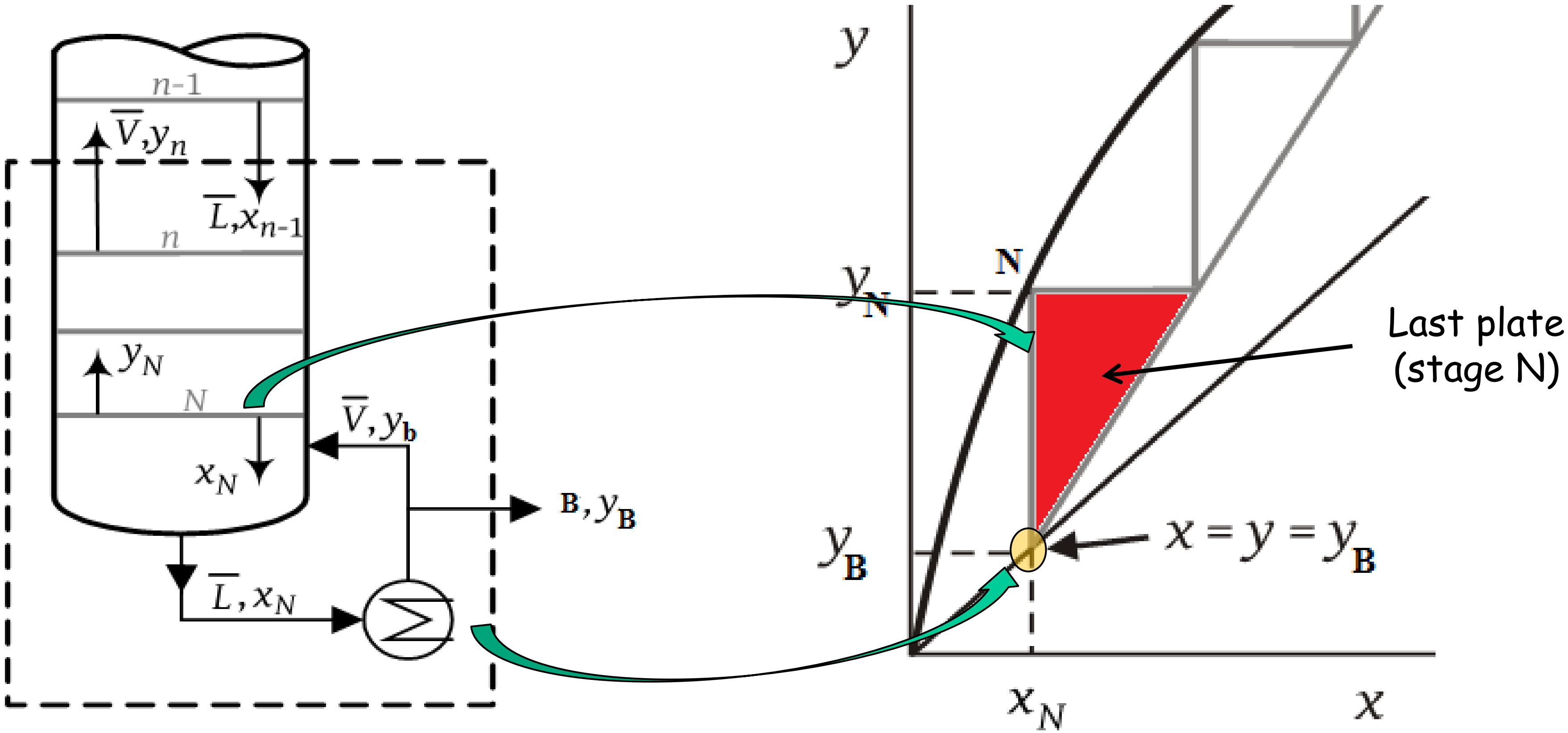


B: Vapor

$$x_N = y_B = y_b$$

Boiler

Total Boiler



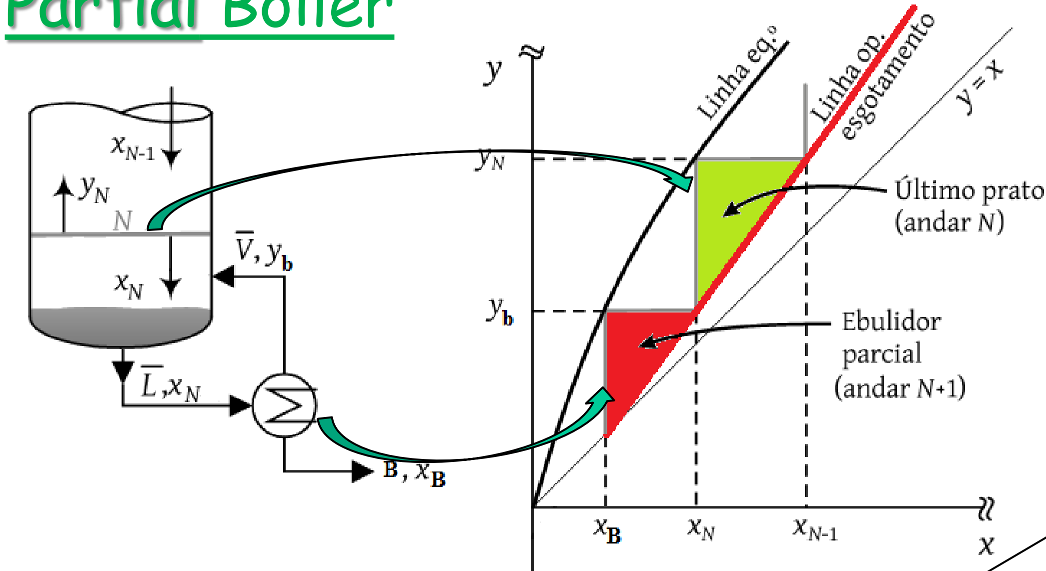
B: Vapor

$$x_N = y_B = y_b$$

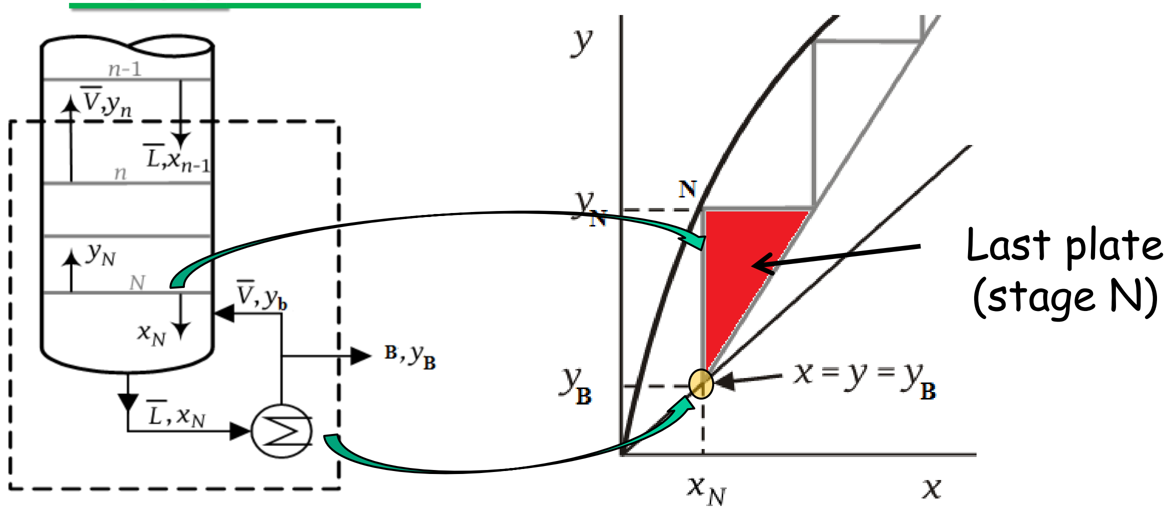
Boiler is not an equilibrium stage

Boiler

Partial Boiler

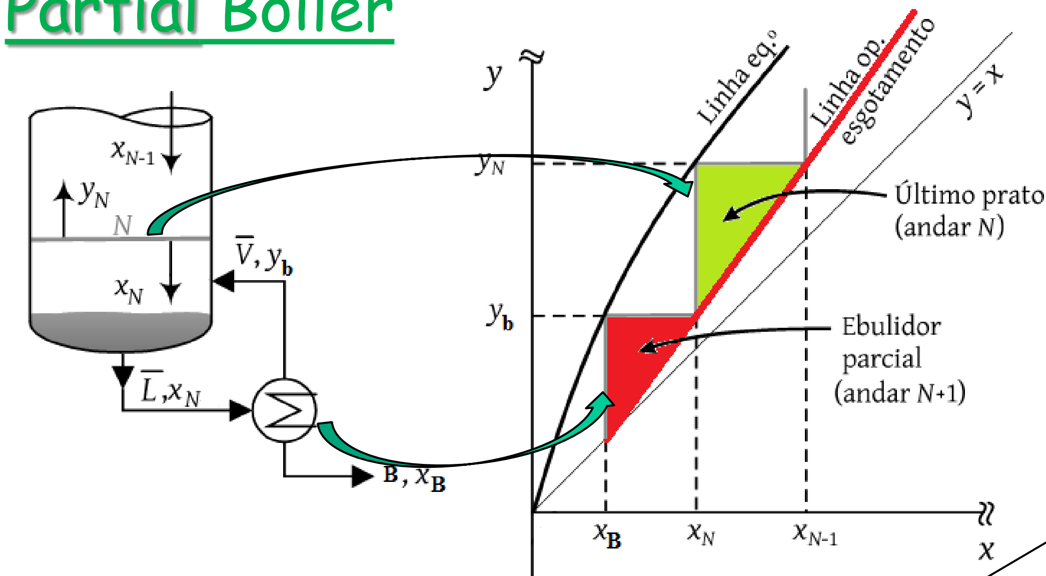


Total Boiler



Boiler

Partial Boiler

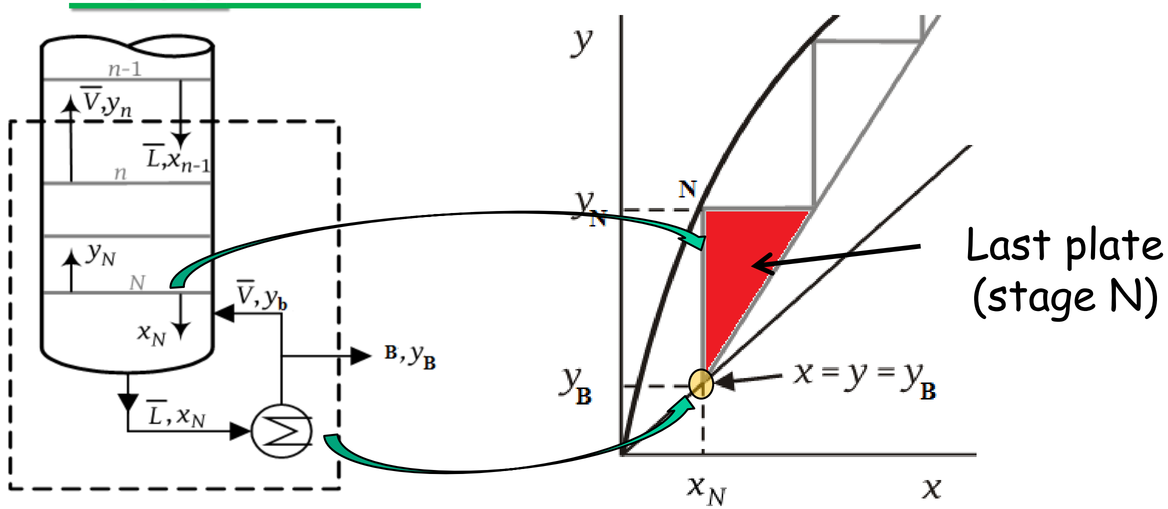


$$N_{column} = N_{MT} - 1$$

Boiler

Total Boiler

$$N_{column} = N_{MT}$$

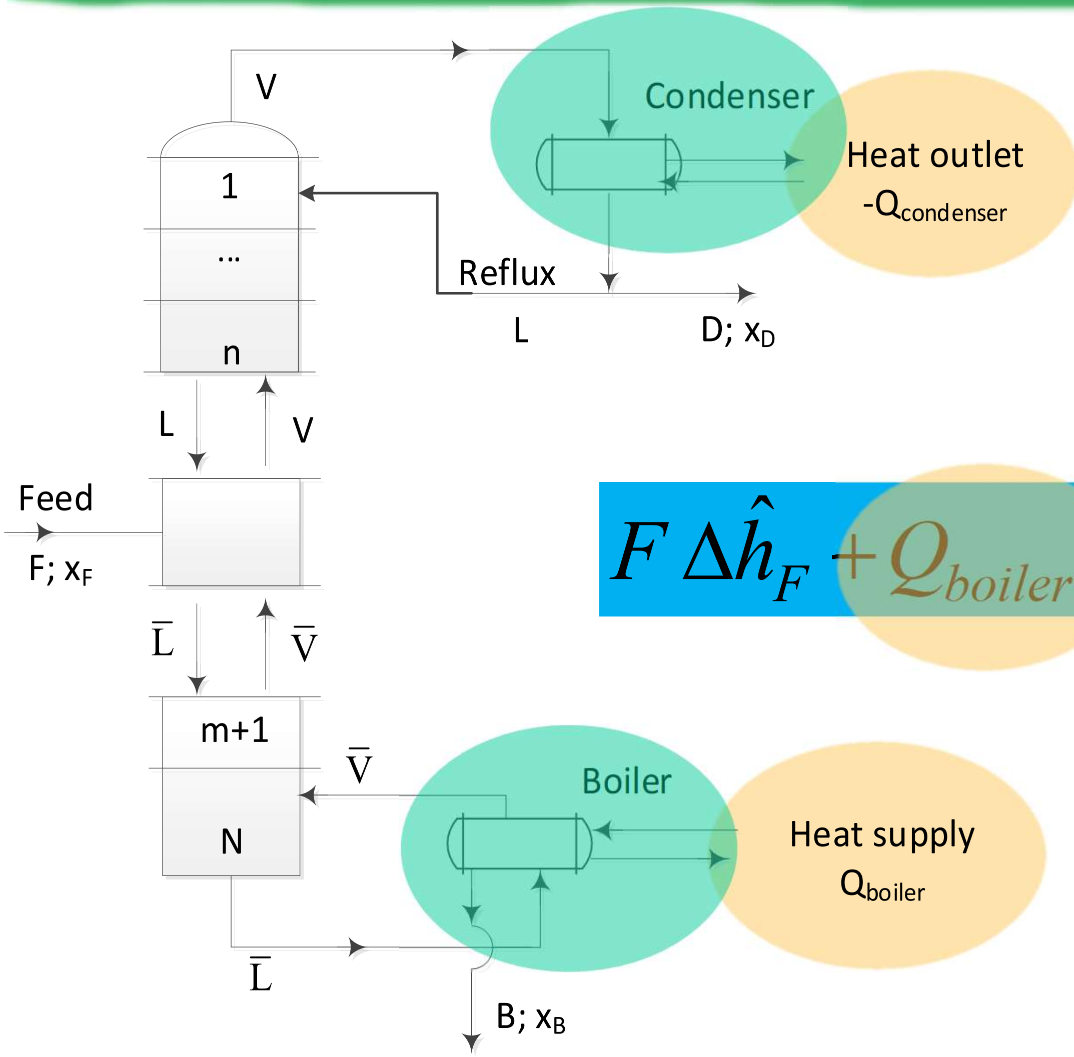


MT: McCabe-Thiele

Summary

- Influence of the reflux ratio on the efficiency of a distillation column
- Optimal reflux ratio
- Condenser and Boiler in a distillation column
- Energy Balances to a distillation column
- Energetic balance to the feed stage
- Stage efficiency

Energy balances

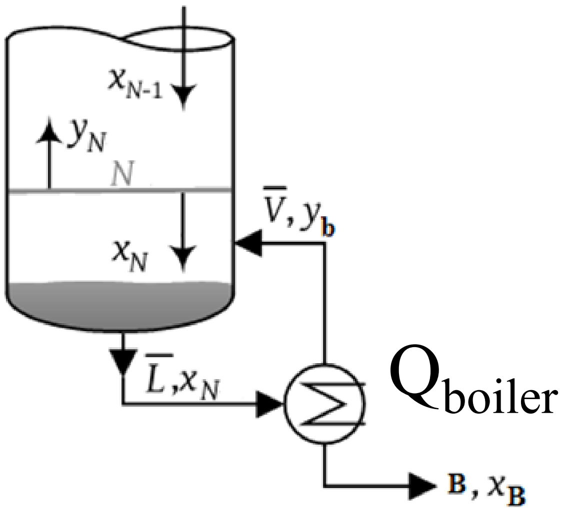


❑ Overall balance

$$F\,\Delta\hat{h}_F + Q_{boiler} = B\,\Delta\hat{h}_B + D\,\Delta\hat{h}_D + Q_{condenser}$$

$\Delta\hat{h}_F$, $\Delta\hat{h}_D$, $\Delta\hat{h}_B$ [J/mol]: specific enthalpies of feed, distillate and residue

Heat required in the boiler, Q_{boiler}



Partial boiler

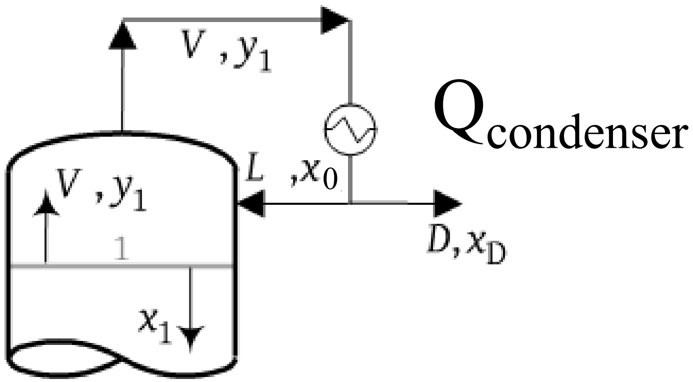
$$\boxed{Q_{boiler} = \bar{V}.\Delta\hat{H}_{vaporization}}$$

$\bar{V}$: molar rate of vapour produced in the boiler

$\Delta\hat{H}_{vap}$: molar enthalpy of vaporization of mixture

(☞ considering sensible heats negligible!)

Heat to be remover from condenser, $Q_{condenser}$



Total Condenser

$$\boxed{Q_{condenser} = V.\Delta\widehat{H}_{condenser}}$$

V: molar rate of vapour fed to the condenser

$\Delta\hat{H}_{cond}$: molar enthalpy of condensation of mixture

Problem 1

A distillation column is used to separate 100 mol/h of a mixture made of two compounds A and C. The feed mixture consists of equal parts of saturated vapor and liquid, with a molar composition of 35% A. It is intended to obtain a distillate with a molar composition of 93% A and a residue with a molar composition of 97.8% C. The reflux ratio is equal to 4.

f) Calculate the heat involved in the boiler and in the condenser.

$$\Delta\hat{H}^{vap}_{(\text{mistura A+C})} = 31.2 \text{ kJ/mol}$$

Heat required in the boiler, Q_{boiler}

$$Q_{boiler} = \bar{V} . \Delta \hat{H}_{vaporization}$$

$\bar{V}$: vapour molar rate produced in the boiler

Heat to be remover from condenser, $Q_{condenser}$

$$Q_{condenser} = V . \Delta \hat{H}_{condenser}$$

V: molar rate of vapour fed to the condenser

Material balances - determine V, L, $\bar{V}$, $\bar{L}$

F = 100 mol/h

D = 36.12 mol/h

B = 63.88 mol/h

$x_F = 0.35$

$x_D = 0.93$

$x_B = 0.022$

R = 4

i = 0.5

Determine L, V
$\bar{L}$, $\bar{V}$

$$R = L/D$$

$$i = \frac{\bar{L} - L}{F}$$

$$V = L + D$$

$$\bar{V} = \bar{L} - B$$

L = 144.5 mol/h
V = 180.6 mol/h

$\bar{L}$ = 194.5 mol/h
$\bar{V}$ = 130.6 mol/h

Heat required in the boiler, Q_{boiler}

$$\boxed{Q_{boiler} = \bar{V} . \Delta\hat{H}_{vaporization}}$$

$\Delta\hat{H}^{vap} = 31.2$ kJ/mol

$\bar{V} = 130.6$ mol/h

$$Q_{caldeira} = 4{,}074.7\ kJ/h$$

Heat to be remover from condenser, $Q_{condenser}$

$$\boxed{Q_{condenser} = V . \Delta\hat{H}_{condenser}}$$

$\Delta\hat{H}^{cond} = -\ 31.2$ kJ/mol

$V = 180.6$ mol/h

$$Q_{condensador} = -5{,}634.7\ kJ/h$$

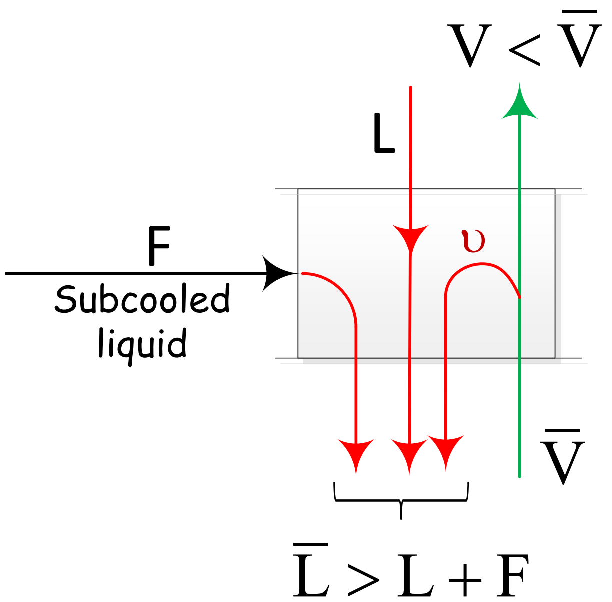
Summary

- Influence of the reflux ratio on the efficiency of a distillation column
- Optimal reflux ratio
- Condenser and Boiler in a distillation column
- Energy Balances to a distillation column
- Energetic balance to the feed stage
- Stage efficiency

Physical state of the Feed

$T_F < T_L$

a) Subcooled liquid



$$\bar{L} = L + F + \upsilon$$

$$\boxed{i = \frac{\bar{L} - L}{F}}$$

$$i = \frac{(L + F + \upsilon) - L}{F}$$

$$i = \frac{F + \upsilon}{F} > 1$$

Value of υ: to be determined by an energy balance to feed plate

Example

A distillation column is used to separate a binary mixture A+C. We know that the feed enters the column at 17 °C. Estimate the slope of the feed operating line.

Feed mixture data:

- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol

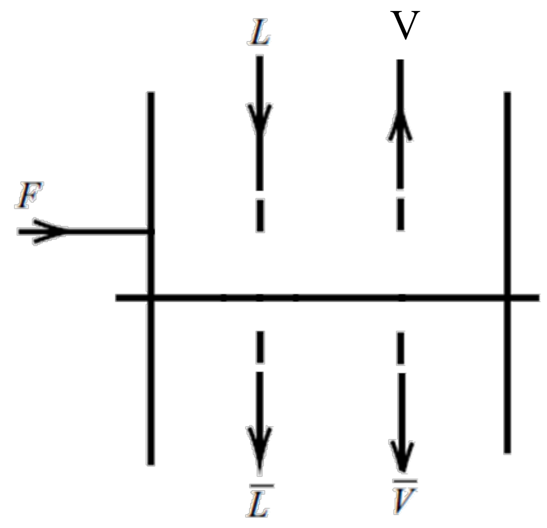
$$\boxed{y_i = \frac{i}{i-1} x_i - \frac{x_F}{i-1}}$$

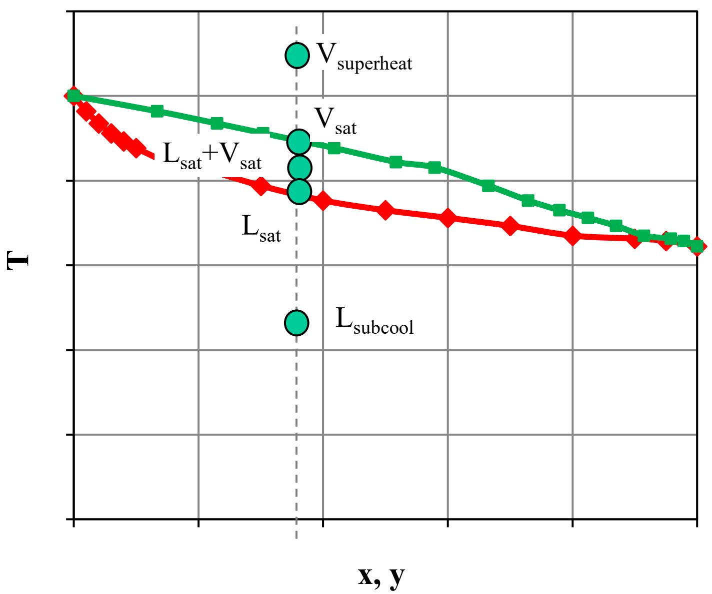
③ Operating line of feed

$$y_i = \frac{i}{i-1}x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol



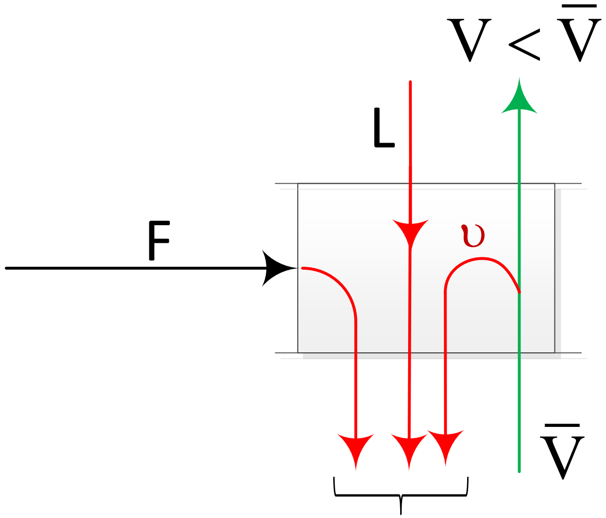


③ Operating line of feed

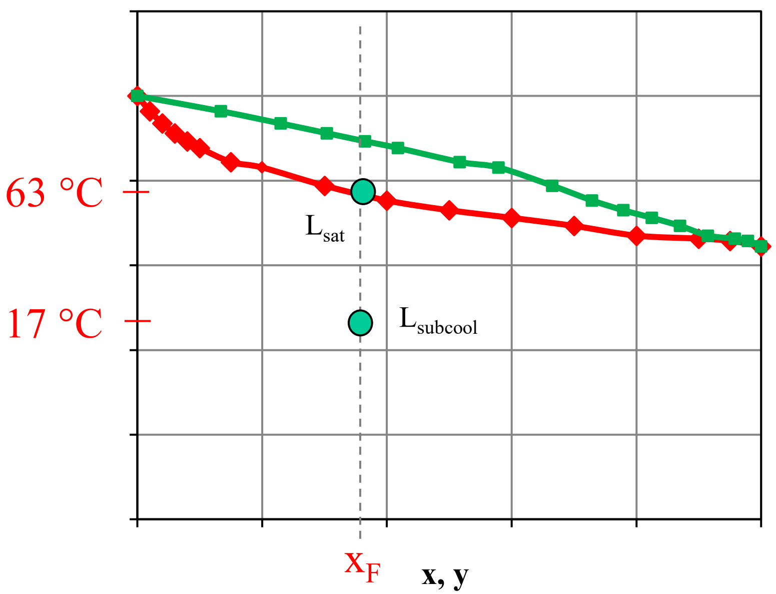
$$y_i = \frac{i}{i-1}x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol



$\bar{L} = L + F + \upsilon$



$T_{feed} < T_{saturated\ liquid}$

Subcooled liquid!

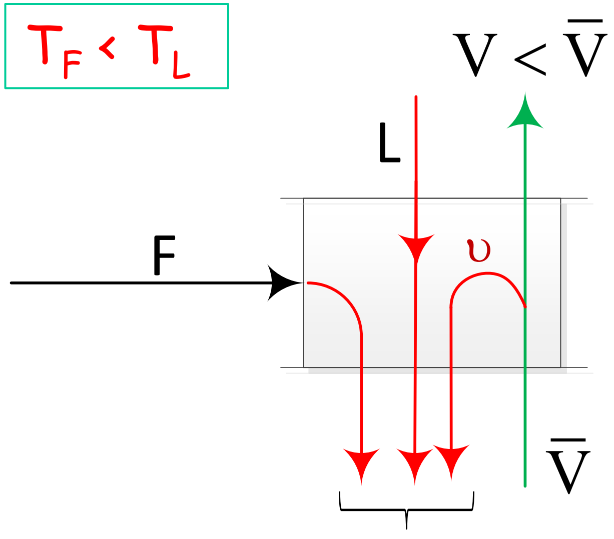
③ Operating line of feed

$$y_i = \frac{i}{i-1} x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol

$T_F < T_L$



$$\bar{L} = L + F + \upsilon$$

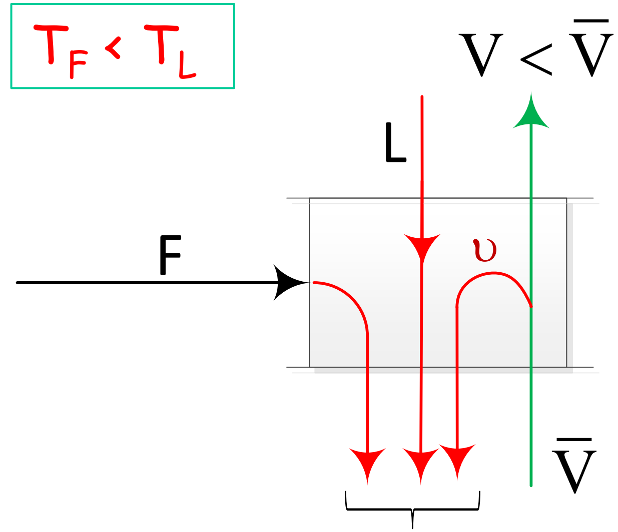
How to estimate υ? ⇒ Energy balance to the feed plate

③ Operating line of feed

$$y_i = \frac{i}{i-1}x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

$T_F < T_L$



$$\bar{L} = L + F + \upsilon$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol

Energy balance to the feed plate

$$\underbrace{F.c_p^{liq}.\Delta T}_{\text{Sensible heat to warm F moles of feed}} = \underbrace{\upsilon.\Delta\widehat{H}_{condensation}}_{\text{Latent heat resulting from the condensation of } \nu \text{ moles of vapor}}$$

$(\Delta T = T_L - T_F)$

$T_{eb,mixture} =$

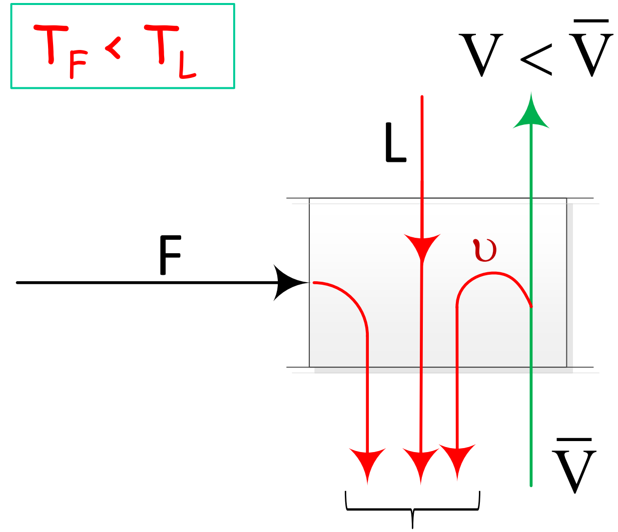
③ Operating line of feed

$$y_i = \frac{i}{i-1} x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol

$T_F < T_L$



$$\bar{L} = L + F + \upsilon$$

Energy balance to the feed plate

$$\underbrace{F.\,c_p^{liq}.\,\Delta T}_{\text{Sensible heat to warm F moles of feed}} = \underbrace{\upsilon.\,\Delta\hat{H}_{condensation}}_{\text{Latent heat resulting from the condensation of } \nu \text{ moles of vapor}}$$

$$\boxed{\upsilon = \frac{F.\,c_p^{liq}.\,\Delta T}{\Delta\hat{H}_{condensation}}}$$

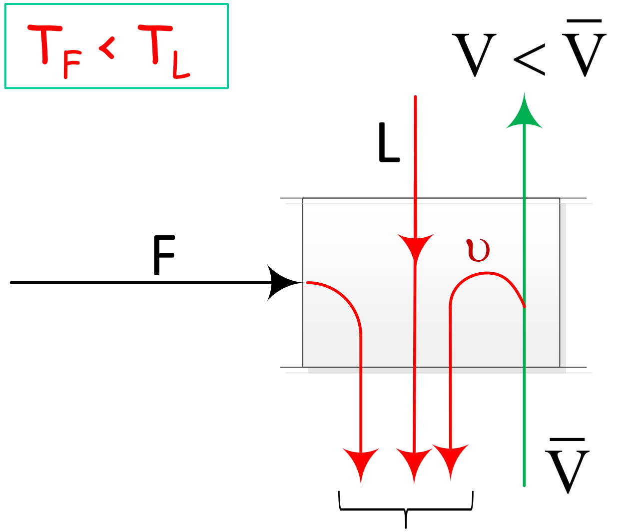
③ Operating line of feed

$$y_i = \frac{i}{i-1} x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L} - L}{F}$$

$T_F < T_L$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol



$$\bar{L} = L + F + \upsilon$$

$$\boxed{\upsilon = \frac{F.c_p^{liq}.\Delta T}{\Delta \hat{H}_{condensation}}}$$

$$\upsilon = \frac{225.4\ Jmol^{-1}K^{-1}\ (336\ K - 290\ K)}{25900\ Jmol^{-1}} F = 0.4\ F$$

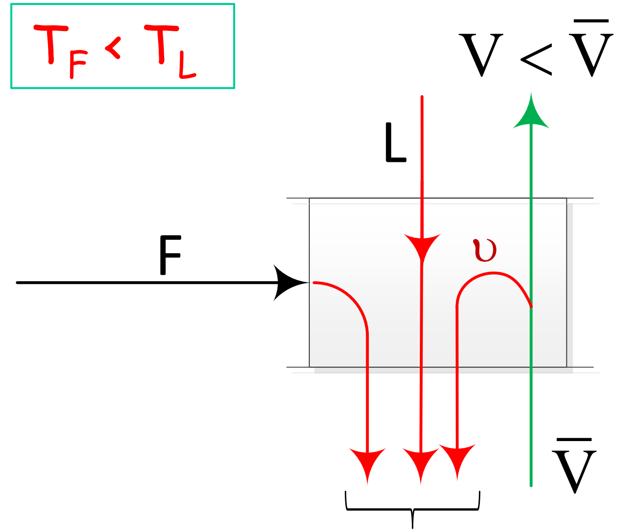
③ Operating line of feed

$$y_i = \frac{i}{i-1}x_i - \frac{x_F}{i-1}$$

$$i = \frac{\bar{L}-L}{F}$$

- Temperature: 17 °C
- Specific heat: 225.4 J/(mol.K)
- Boiling point: 63 °C
- Enthalpy of vaporization: 25 900 J/mol

$T_F < T_L$

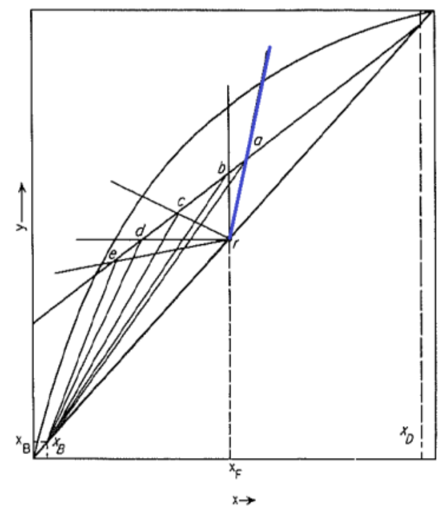


$$\bar{L} = L + F + \upsilon$$

$$\upsilon = 0.4F$$

$$i = \frac{\bar{L}-L}{F} = \frac{(L+F+0.4F)-L}{F} = \mathbf{1.4}$$

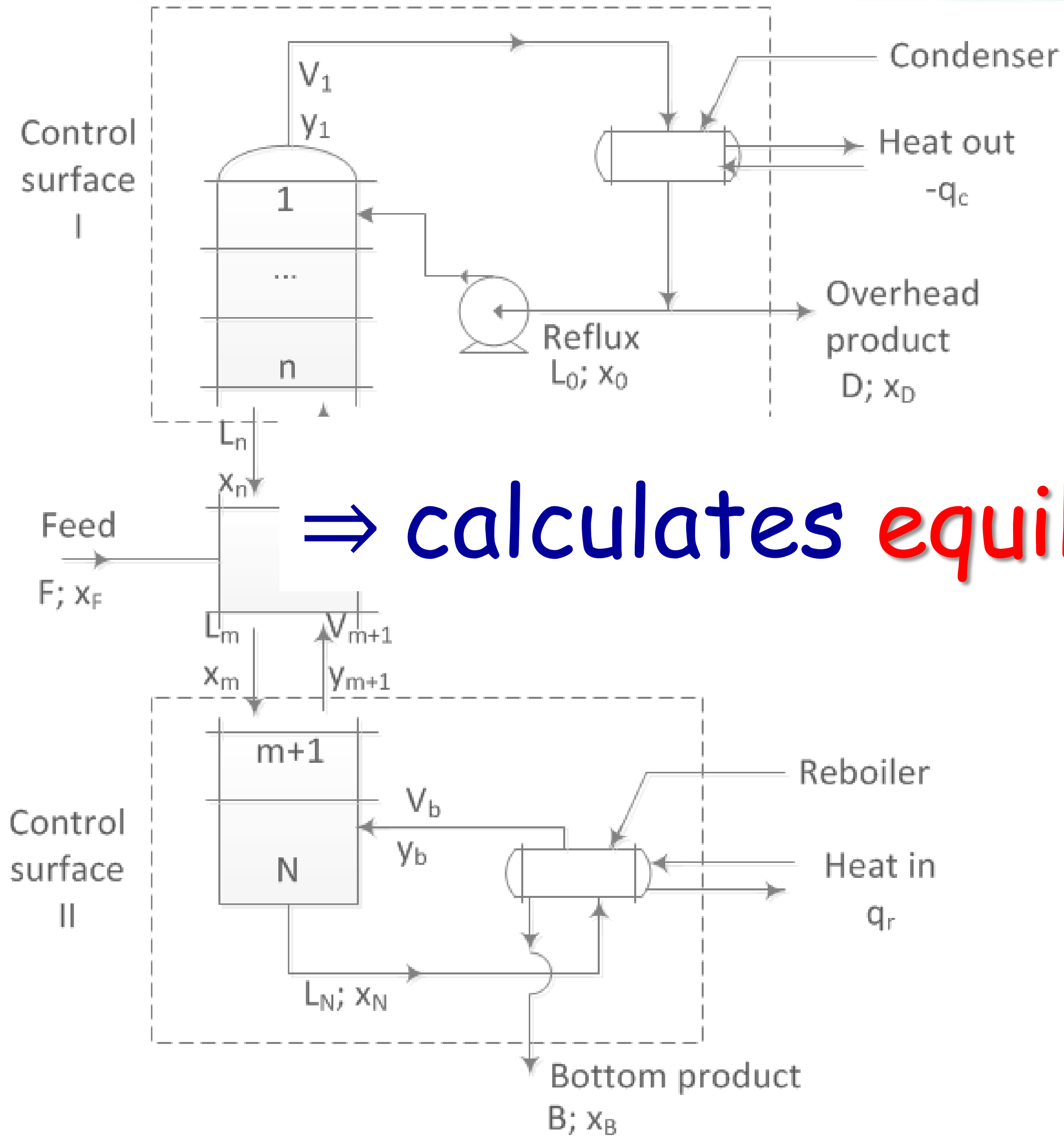
$$y_i = \frac{i}{i-1}x_i - \frac{x_F}{i-1} = \mathbf{3.5}\boldsymbol{x_i} - \mathbf{0.685}$$

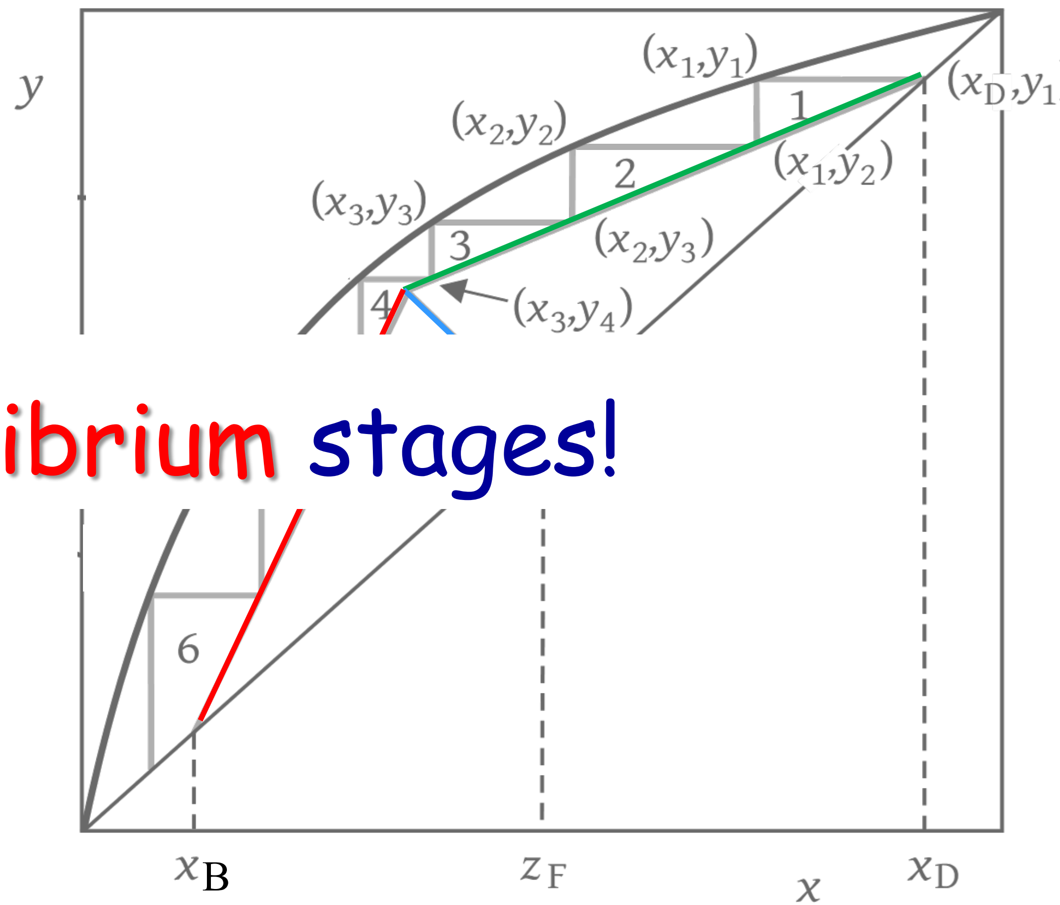


Summary

- Influence of the reflux ratio on the efficiency of a distillation column
- Optimal reflux ratio
- Condenser and Boiler in a distillation column
- Energy Balances to a distillation column
- Energetic balance to the feed stage
- Stage efficiency

McCabe – Thiele Method





⇒ calculates equilibrium stages!

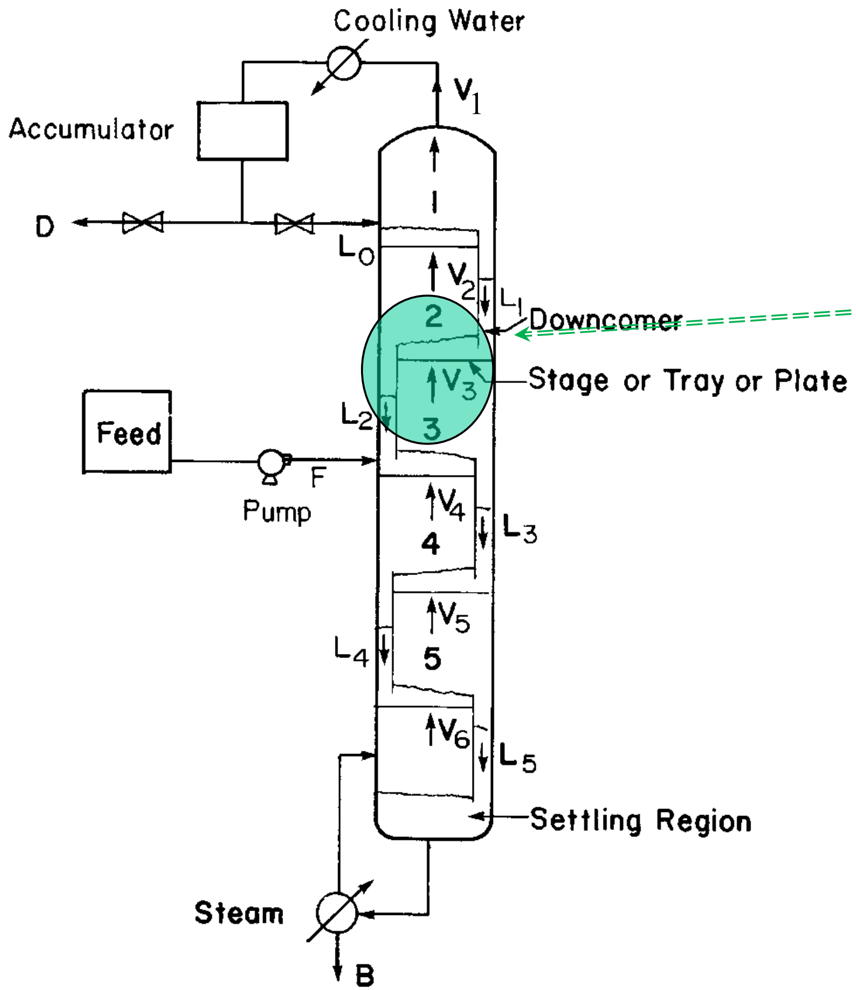
Plate efficiency

In practice, a real plate is **not** an equilibrium plate

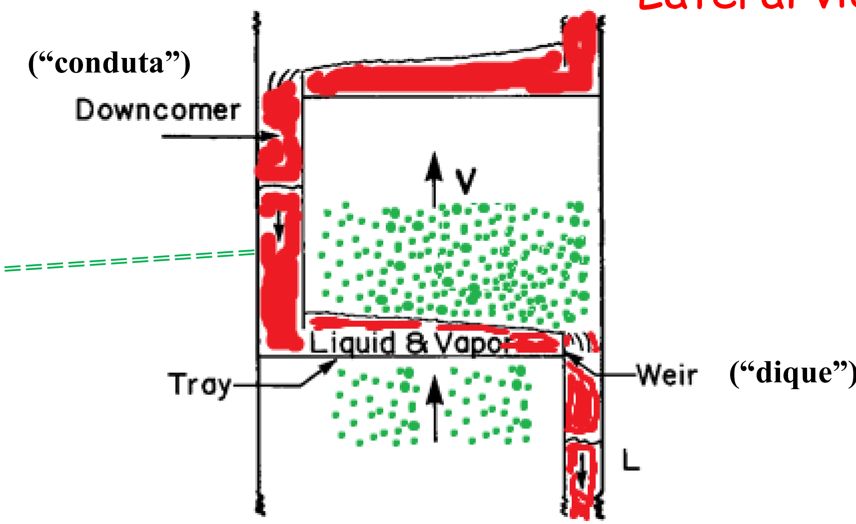
Due to, among others...

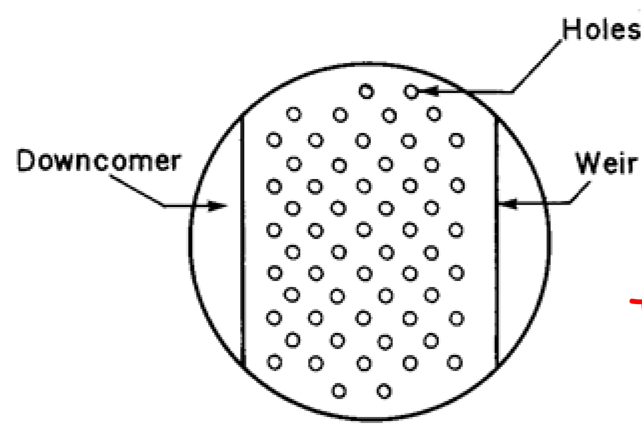
- Insufficient time of contact between the two phases
- Deficient mixing of the two phases in the plate

$\Rightarrow$ The liquid and vapour streams exiting the plate are not in equilibrium!



Lateral view





Top view

https://www.youtube.com/watch?v=BaBMXgVBQKk&t=2s

Total efficiency (of the column…)

$$E_0 = \frac{N_{\text{equilibrium stages}}}{N_{\text{real stages}}} \leq 1$$

Depends on:

- Physical properties of currents L and V (ρ, μ)
- Flow conditions (V/L)
- Mass and heat transfer (Re, Pr, Sc, Nu, Sh)

<u>It is determined by</u>:

- comparison with performance data from industrial columns
- use of empirical efficiency equations derived from data on industrial columns
- use of semi theoretical models based on mass- and heat-transfer rates
- scale-up from laboratory or pilot plant columns

Total efficiency (of the column...)

$$\boxed{E_0 = \frac{N_{\text{equilibrium stages}}}{N_{\text{real stages}}}} \leq 1$$

For distillation processes where the compounds are close to their boiling points and the viscosity of the liquid phases is low, the efficiency of tray columns is often over 70% and can even reach 100%

- **Energetic efficiency on distillation columns**

Energetic efficiency

Distillation – intensive energy operation

Optimization of distillation => ◇ optimize operating conditions
◇ minimize energy consumption

Practical example: R ↗ =>
- Energy consumption ↗
- N_{stages} ↘

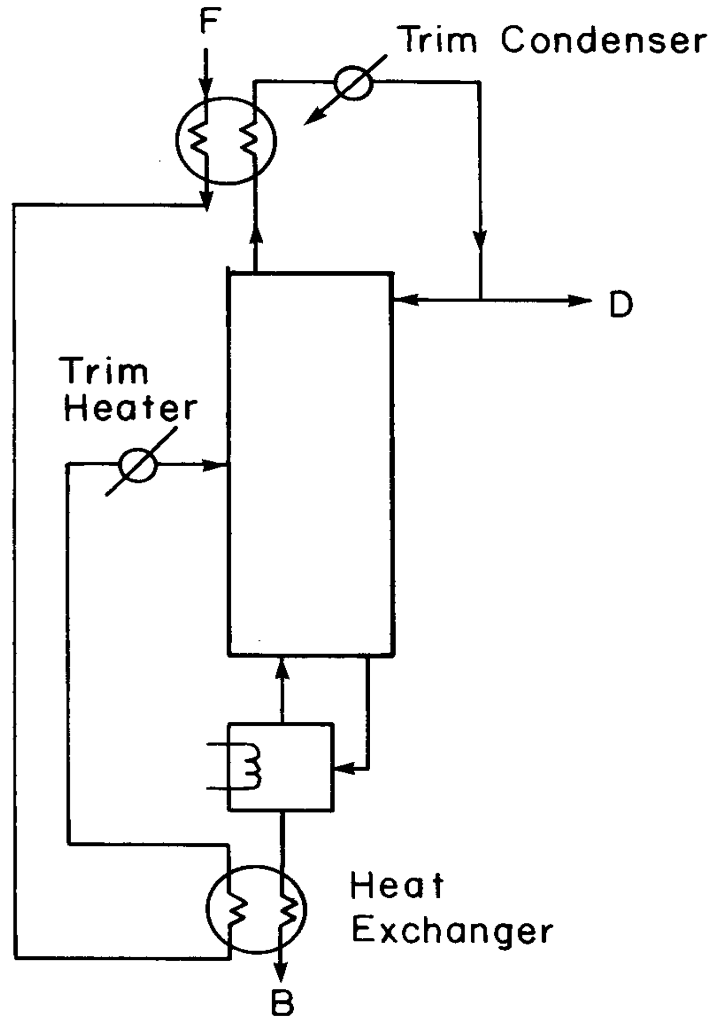
That is, a compromise must be established between **equipment costs** (i.e. sieve trays) and **operational costs** (i.e. energy costs)!

Energetic efficiency

How to optimize the energy demanding of a distillation unit?

Energetic integration

Hot streams are used to warm cold streams



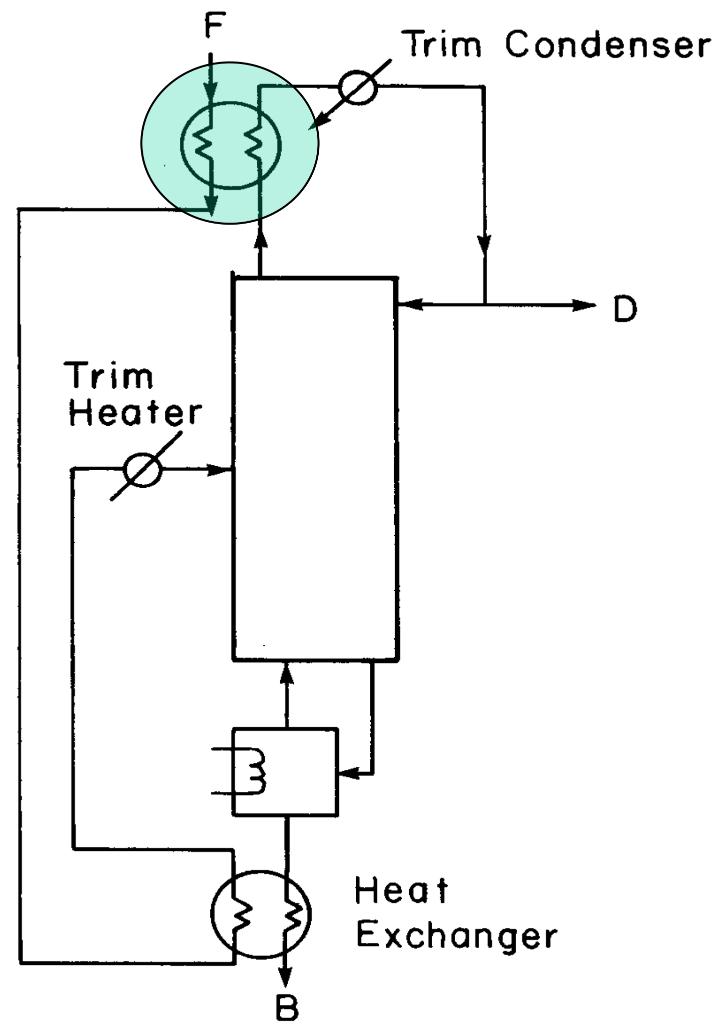
From *Separation Process Engineering, Third Edition* by Phillip C. Wankat
(ISBN: 0131382276)

Energetic integration

Hot streams are used to warm **cold streams**

Examples:

- Feed is pre-heated in the condenser where the exiting vapor condenses in contact with the cold feed, that in turn will heat



From *Separation Process Engineering, Third Edition* by Phillip C. Wankat

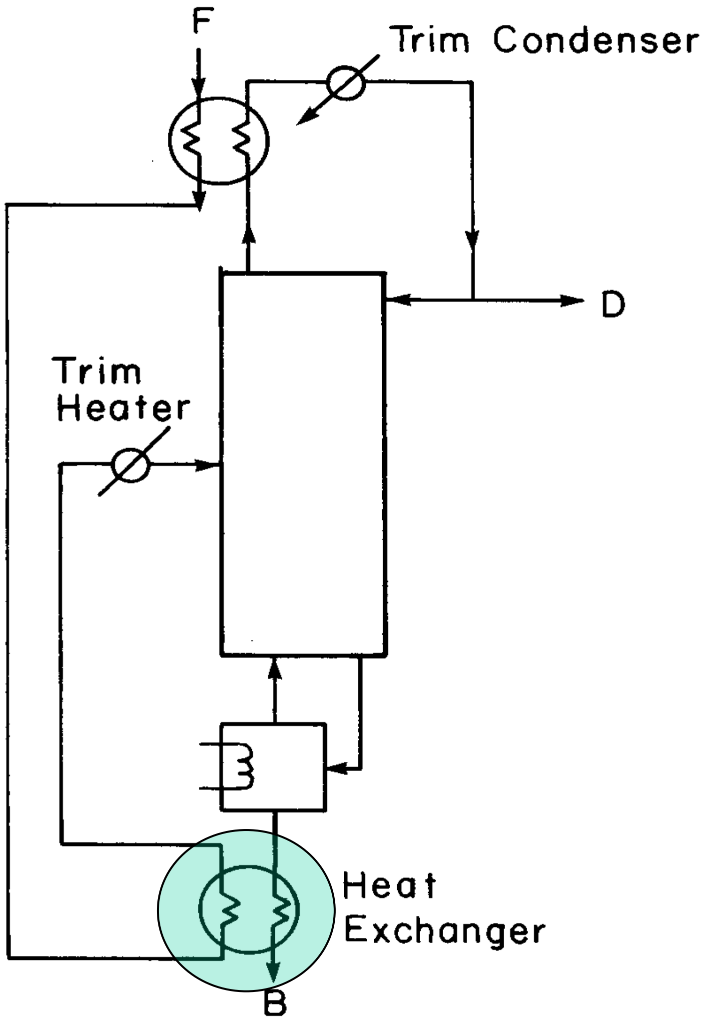
Energetic integration

Hot streams are used to warm cold streams

Examples:

- Feed is pre-heated in the condenser where the exiting vapor condenses in contact with the cold feed, that in turn will heat

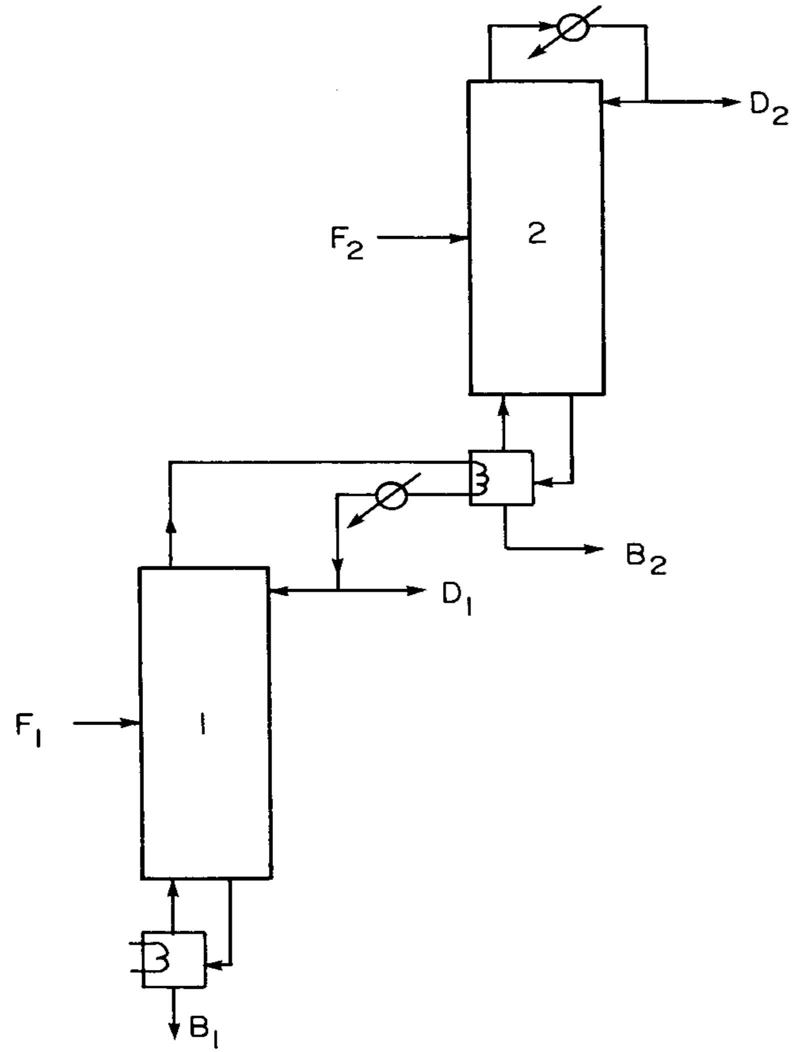
- Feed is pre-heated by the bottom current, that acts as a heating fluid



From *Separation Process Engineering, Third Edition* by Phillip C. Wankat

Integrated several distillation columns

The vapour of column 1 is condensed in the boiler of column 2



1.05 a $1.5 \times R_{min}$

Table 7.1 Representative Commercial Binary Distillation Operations

Binary Mixture	Average Relative Volatility	Number of Trays	Typical Operating Pressure, psia	Reflux-to-Minimum-Reflux Ratio
1,3-Butadiene/vinyl acetylene	1.16	130	75	1.70
Vinyl acetate/ethyl acetate	1.16	90	15	1.15
o-Xylene/*m*-xylene	1.17	130	15	1.12
Isopentane/*n*-pentane	1.30	120	30	1.20
Isobutane/*n*-butane	1.35	100	100	1.15
Ethylbenzene/styrene	1.38	34	1	1.71
Propylene/propane	1.40	138	280	1.06
Methanol/ethanol	1.44	75	15	1.20
Water/acetic acid	1.83	40	15	1.35
Ethylene/ethane	1.87	73	230	1.07
Acetic acid/acetic anhydride	2.02	50	15	1.13
Toluene/ethylbenzene	2.15	28	15	1.20
Propane/1,3-butadiene	2.18	40	120	1.13
Ethanol azeotrope/water	2.21	60	15	1.35
Isopropanol/water	2.23	12	15	1.28
Benzene/toluene	3.09	34	15	1.15
Methanol/water	3.27	60	45	1.31
Cumene/phenol	3.76	38	1	1.21
Benzene/ethylbenzene	6.79	20	15	1.14
HCN/water	11.20	15	50	1.36
Ethylene oxide/water	12.68	50	50	1.19
Formaldehyde/methanol	16.70	23	50	1.17
Water/ethylene glycol	81.20	16	4	1.20

Mix, T.W., J.S. Dweck, M. Weinberg, and R.C. Armstrong, Chem. Eng. Prog., 74(4), 49–55 (1978).