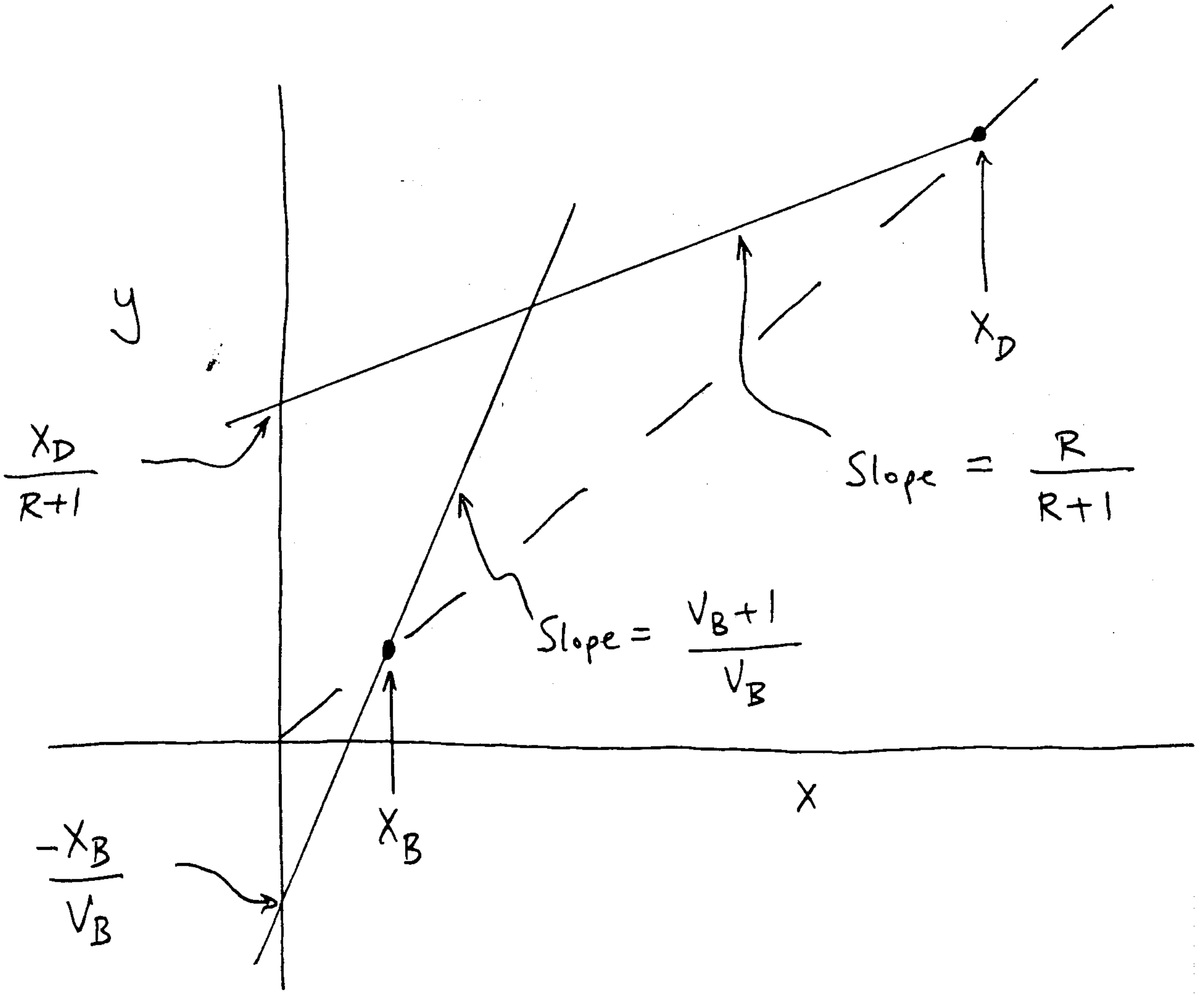
LET'S LOOK AT SOME SIMILARITIES

q-line: $$y = \left(\frac{\psi-1}{\psi}\right) x + \frac{1}{\psi} z$$

RECTIFIER OP. LINE: $$y_{n+1} = \left(\frac{R}{R+1}\right) x_n + \left(\frac{1}{R+1}\right) x_D$$

STRIPPER OP. LINE: $$y_m = \left(\frac{V_B+1}{V_B}\right)(x_{m-1}) - \frac{1}{V_B} x_B$$



EXAMPLE: Look at similar problem to Rectifier Example

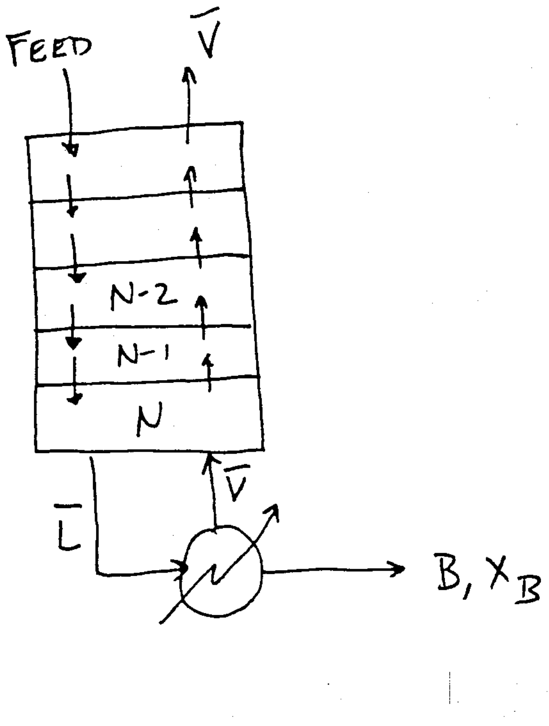
Hexane (1) / octane (2)

- Spec is 95% octane
- Partial reboiler
- Boilup ratio = 2
- Uniform pressure of 1 bar
- XY phase diagram given
- FEED 100 mol/hr (all liquid)
- STRIPPER

⇒ HOW GOOD CAN WE DO?

SOLUTION:

STEP 1: PICTURE



STEP 2: WHAT DO WE KNOW?

→ $X_B = 0.05$ (95% is octane)

→ Reboiler is an EQ Stage.

→ $F = 100$ mol/hr

→ $V_B = \frac{\bar{V}}{B} = 2$

→ Pressure = 1 bar everywhere.

STEP 3: BULK FLOWS

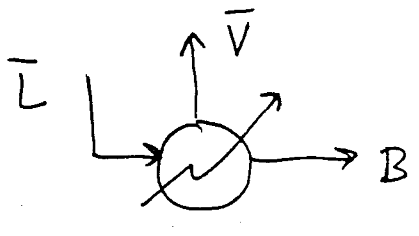
$$F = \bar{V} + B$$

$$\boxed{100 = \bar{V} + B}$$ (overall mole balance)

$$V_B = 2 = \frac{\bar{V}}{B} \Rightarrow \boxed{\bar{V} = 2B}$$ (boilup ratio)

$$100 = 2B + B \Rightarrow \boxed{\begin{array}{l} B = 33.3 \text{ mol/hr} \\ \bar{V} = 66.7 \text{ mol/hr} \end{array}}$$

STEP 4: BALANCE AROUND REBOILER



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

$$\Rightarrow \boxed{L = 100 \text{ mol/hr}}$$

STEP 5: Operating Line

Slope: $\frac{V_B+1}{V_B} = \frac{2+1}{2} = \frac{3}{2} = 1.5$

(Note: Double check slope > 1 for stripper)

$x_B = 0.05$ (starting point on 45° line)

STEP 6: ONTO XY DIAGRAM

(see next page)

STEP 7: STEP OFF STAGES

① $x_B = 0.05$ (spec)

② Streams leaving stage in EQ

$\Rightarrow$ x_B + y_{N+1} in EQ (EQ Line)

$y_{N+1} = 0.2$

③ y_{N+1} + x_N are passing streams (OP. Line)

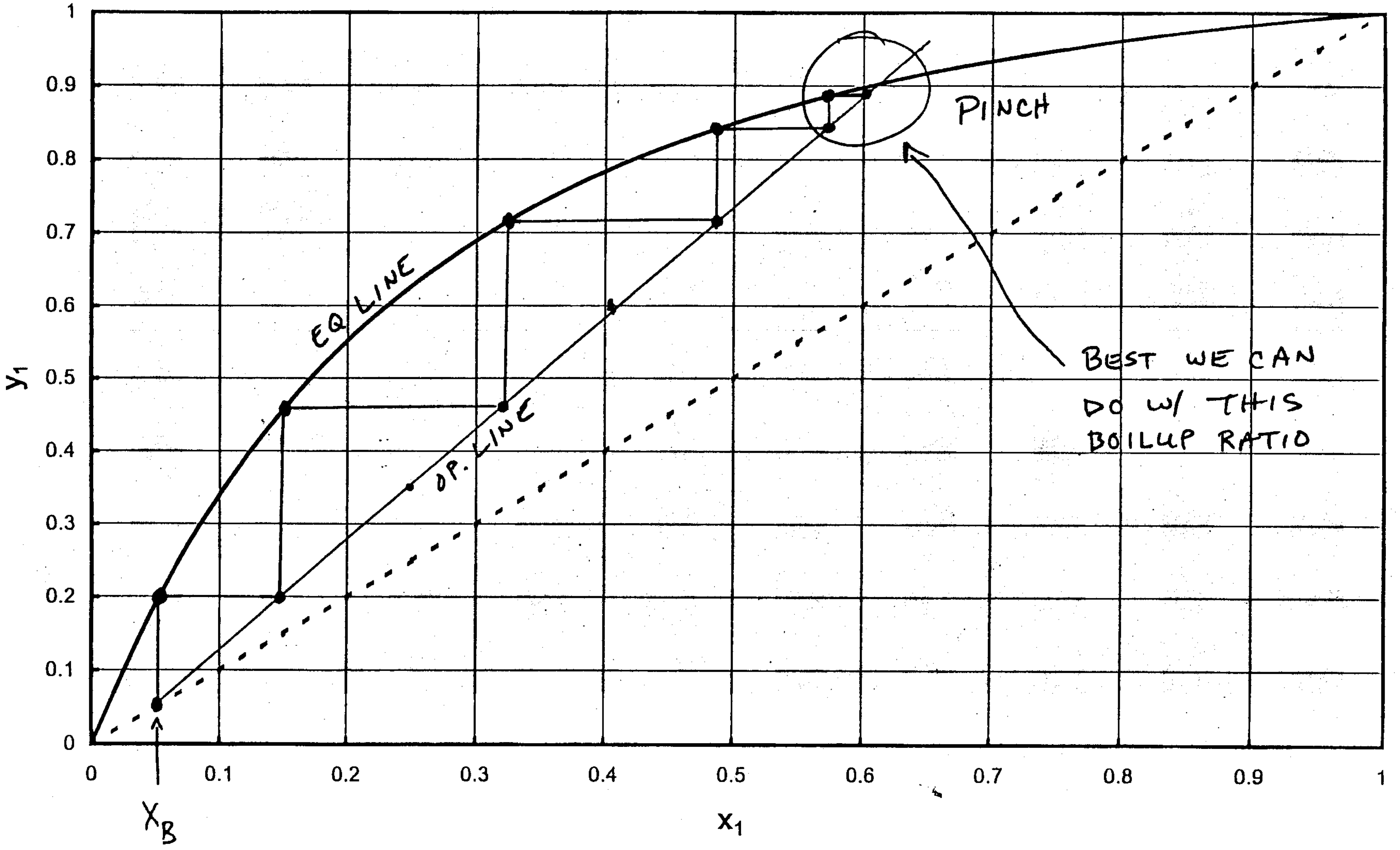
$x_N = 0.15$

④ x_N + y_N are in EQ. (EQ. Line)

$y_N = 0.46$

STRIPPER ONLY

XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar



⑤ y_N + x_{N-1} are passing streams (OP. Line)

$x_{N-1} = 0.325$

⋮ and so on.

WHAT HAPPENS?

EVENTUALLY, EQ Line + OP Line touch ⇒ best we can do (with given V_B)

⇒ Design Change.

BOILUP RATIO

a) V_B smaller

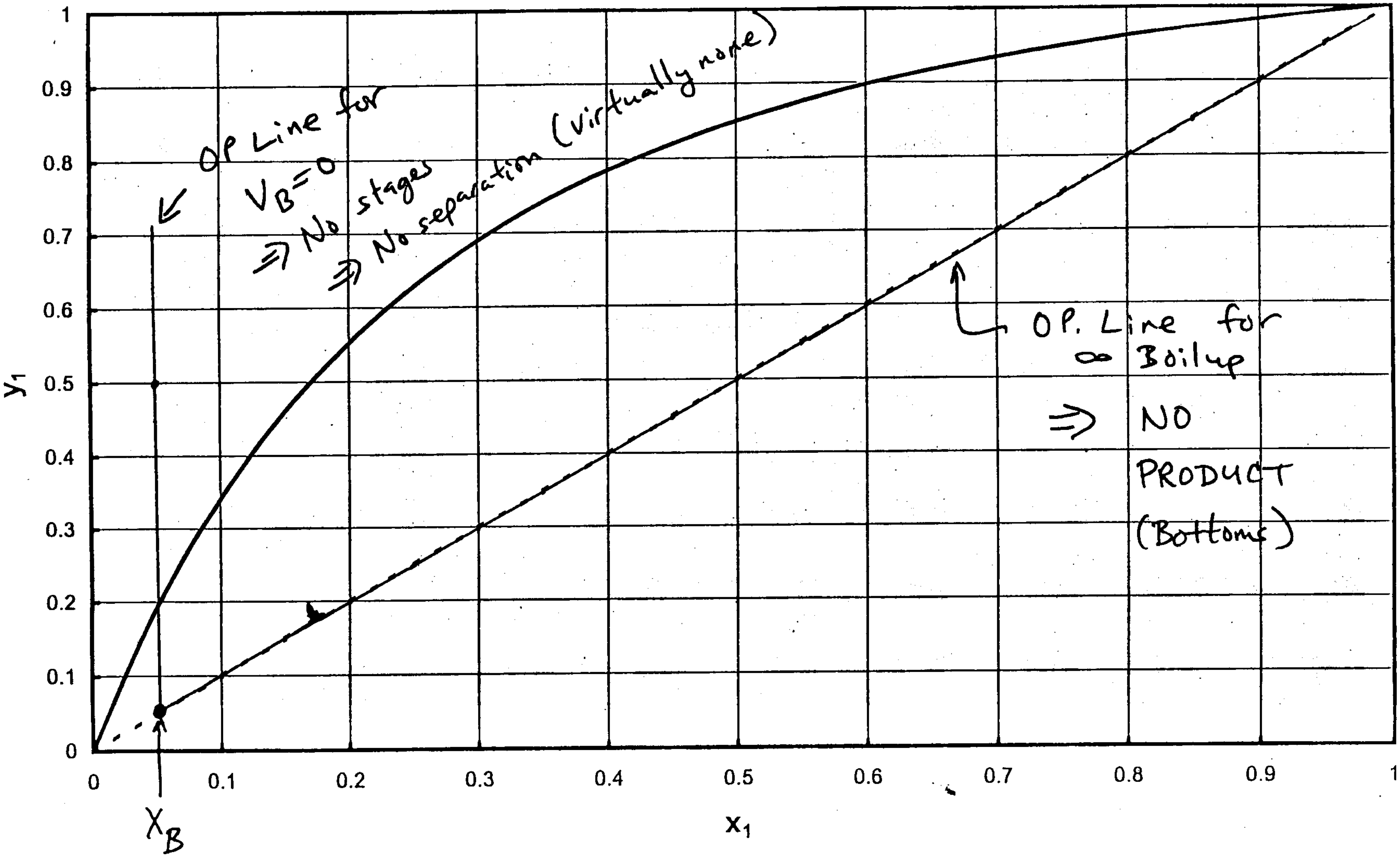
$$\frac{V_B+1}{V_B} \rightarrow \infty$$

b) V_B larger

$$\frac{V_B+1}{V_B} \rightarrow 1$$

[redacted]

XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar



NEITHER OPTION IS GOOD

⇒ Very similar to rectifier.

ALSO, PINCH AT END POINT (∞ Stages)

SOLUTION?

⇒ Transition from one Operating line (Rectifier) to the other (Stripper)

(See attached)

Next time . . .

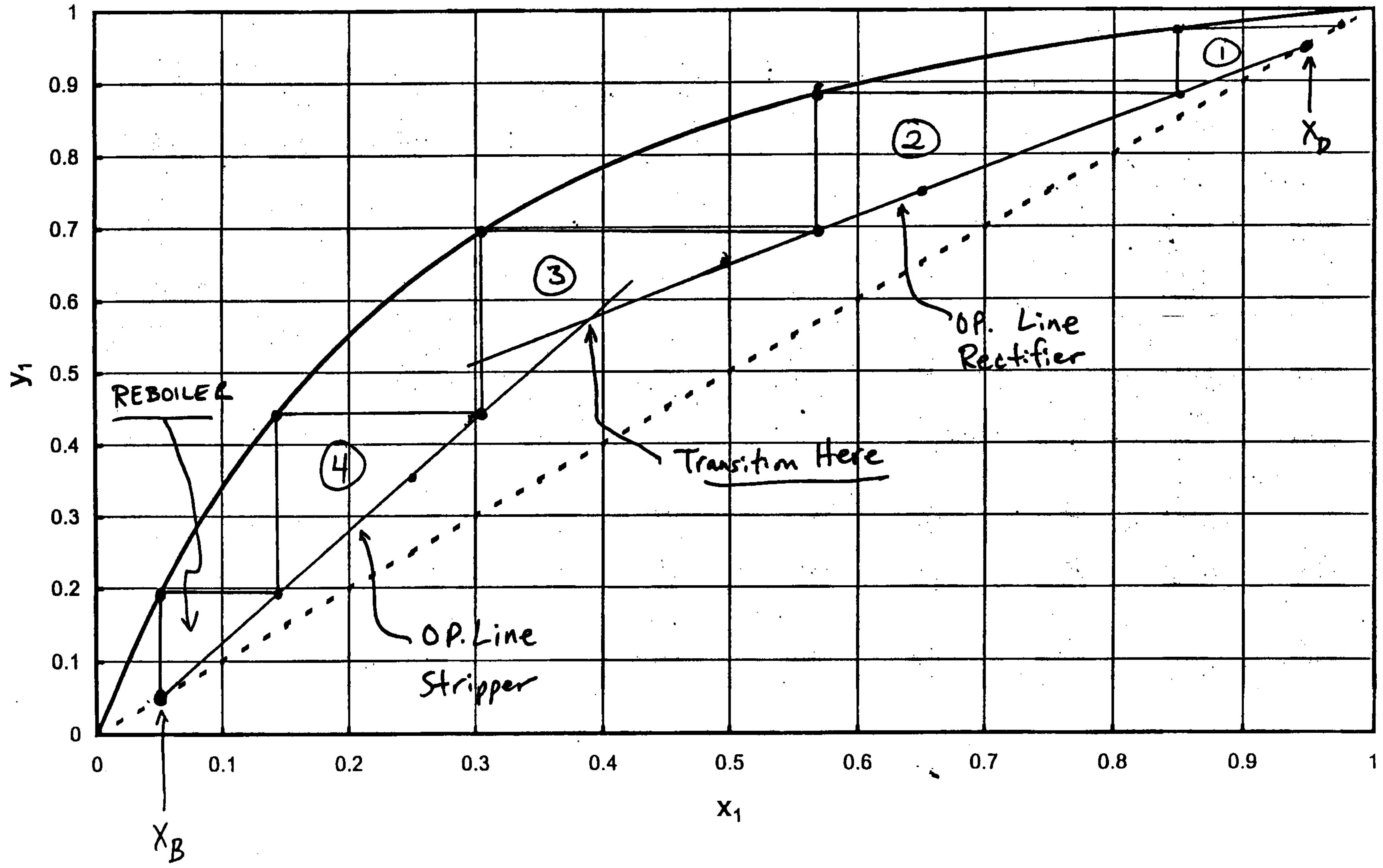
L + $\bar{L}$

V + $\bar{V}$

and the Feed Stream.

DISTILLATION COLUMN

XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar



COLUMN DESIGN:

A. FEED CONDITION

Ψ

POSSIBILITIES

i) Subcooled Liquid (SCL)

ii) Saturated Liquid (SL)

iii) Liquid + Vapor (L+V)

iv) Saturated Vapor (SV)

v) Superheated Vapor (SHV)

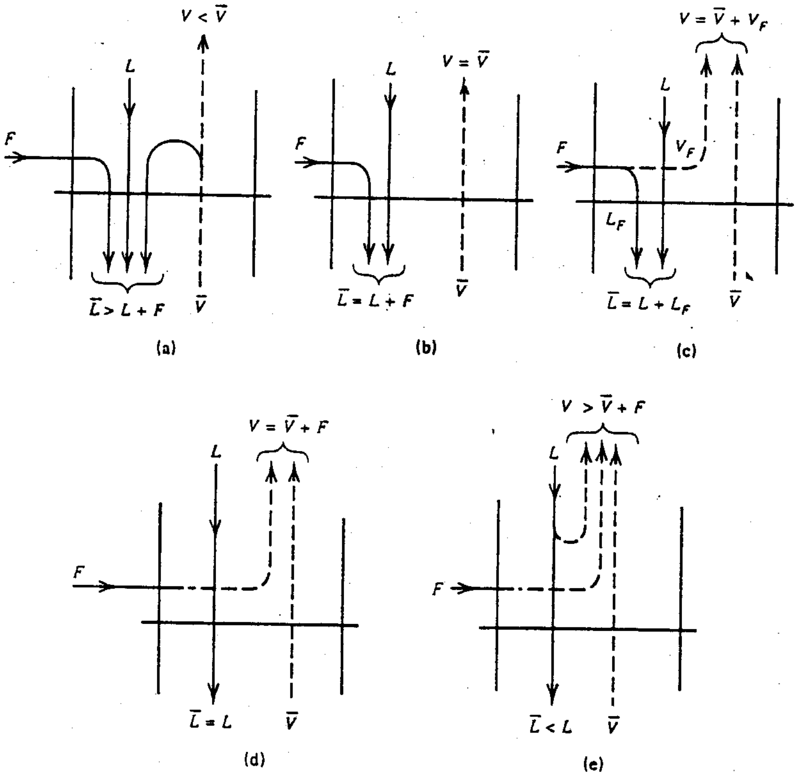


Figure 7.7 Possible feed conditions: (a) subcooled liquid feed; (b) bubble-point liquid feed; (c) partially vaporized feed; (d) dew-point vapor feed; (e) superheated vapor feed. [Adapted from W.L. McCabe, J.C. Smith, and P. Harriott, *Unit Operations of Chemical Engineering*, 5th ed., McGraw-Hill, New York (1993).]

FiG. 7.7

SEADER
+
HENLEY
1998

RECALL q-line:

$$\left.\Psi = \frac{V}{F}\ \right\}\ \text{Flash} \Rightarrow$$

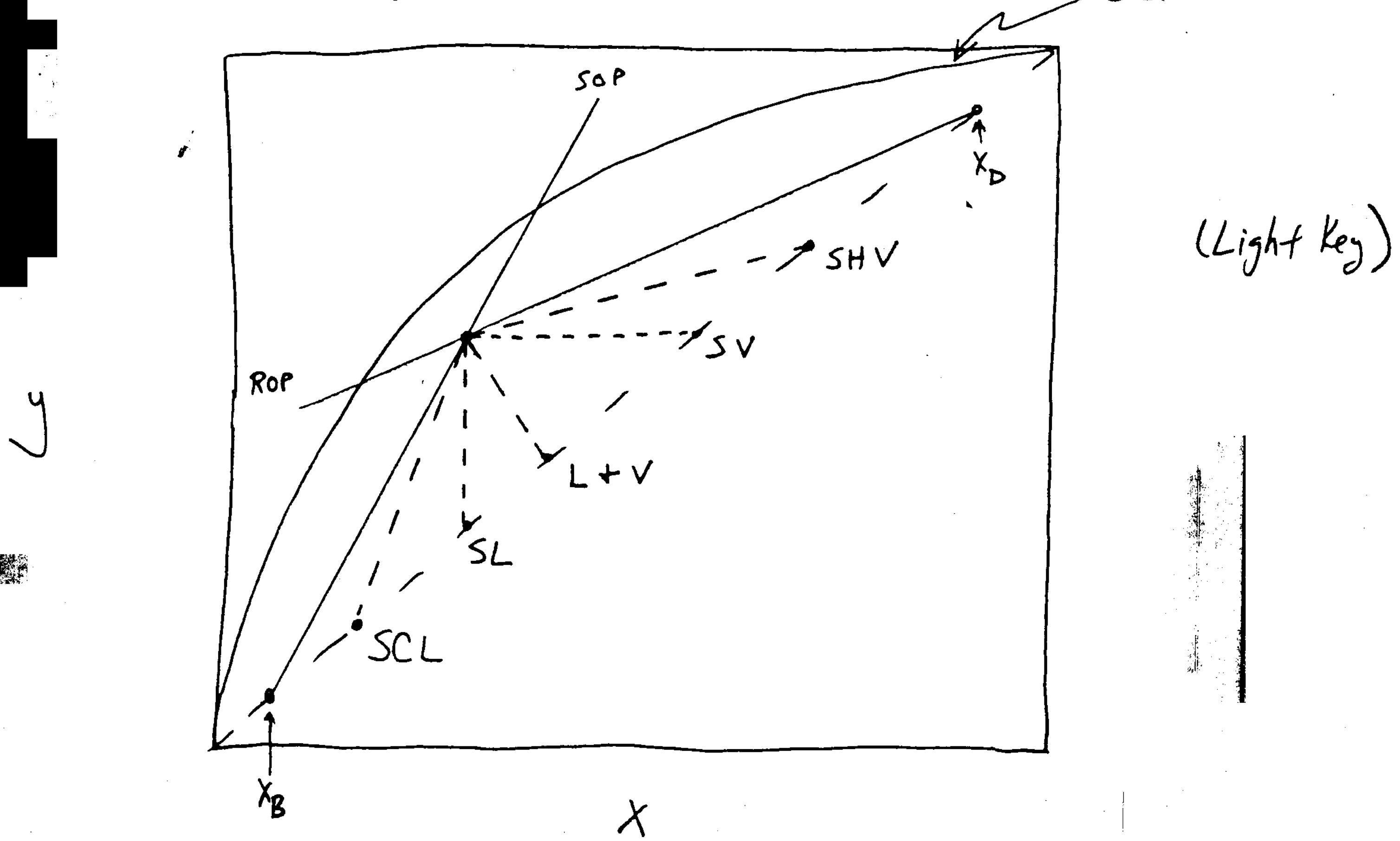
V is amount (moles) of Feed stream vaporized.

(NOT VAPOR FLOW RATE IN RECTIFIER)

Equation q-line:

$$\underset{\substack{\uparrow \\ \text{VAPOR} \\ \text{COMPOSITION}}}{y} = \left(\frac{\Psi-1}{\Psi}\right)\underset{\substack{\uparrow \\ \text{Liquid} \\ \text{COMPOSITION}}}{x} + \left(\frac{1}{\Psi}\right)\underset{\substack{\uparrow \\ \text{FEED} \\ \text{COMPOSITION}}}{z}$$

XY PHASE DIAGRAM



FROM PREVIOUS FIGURE . . .

KNEW

SOP (STRIPPER OPERATING LINE)

ROP (RECTIFIER OPERATING LINE)

INTERSECTION GIVES <u>ONE POINT</u> ON q-LINE

IF YOU KNOW THE CONDITION OF THE FEED, CAN DRAW q-LINE + GET FEED COMPOSITION.

$\psi \quad \left(\text{Slope} = \frac{\psi - 1}{\psi} \right)$

NOTE: FOR SUBCOOLED LIQUID

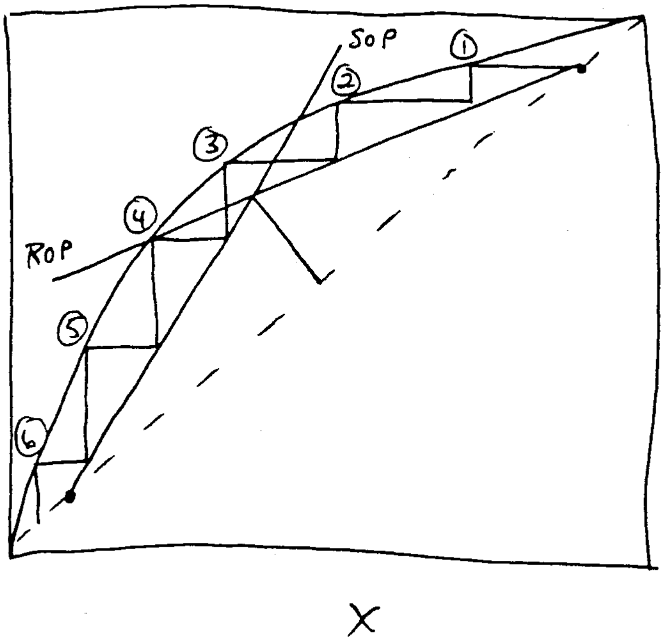
+ SUPERHEATED VAPOR

⇒ ENERGY BALANCE

(SEE BOOK p. 368, SEADER + HENLEY)

FEED STAGE LOCATION:

⇒ TELLS ME WHEN TO SWITCH FROM ROP to SOP.



FEED AT STAGE 3 (OPTIMAL)

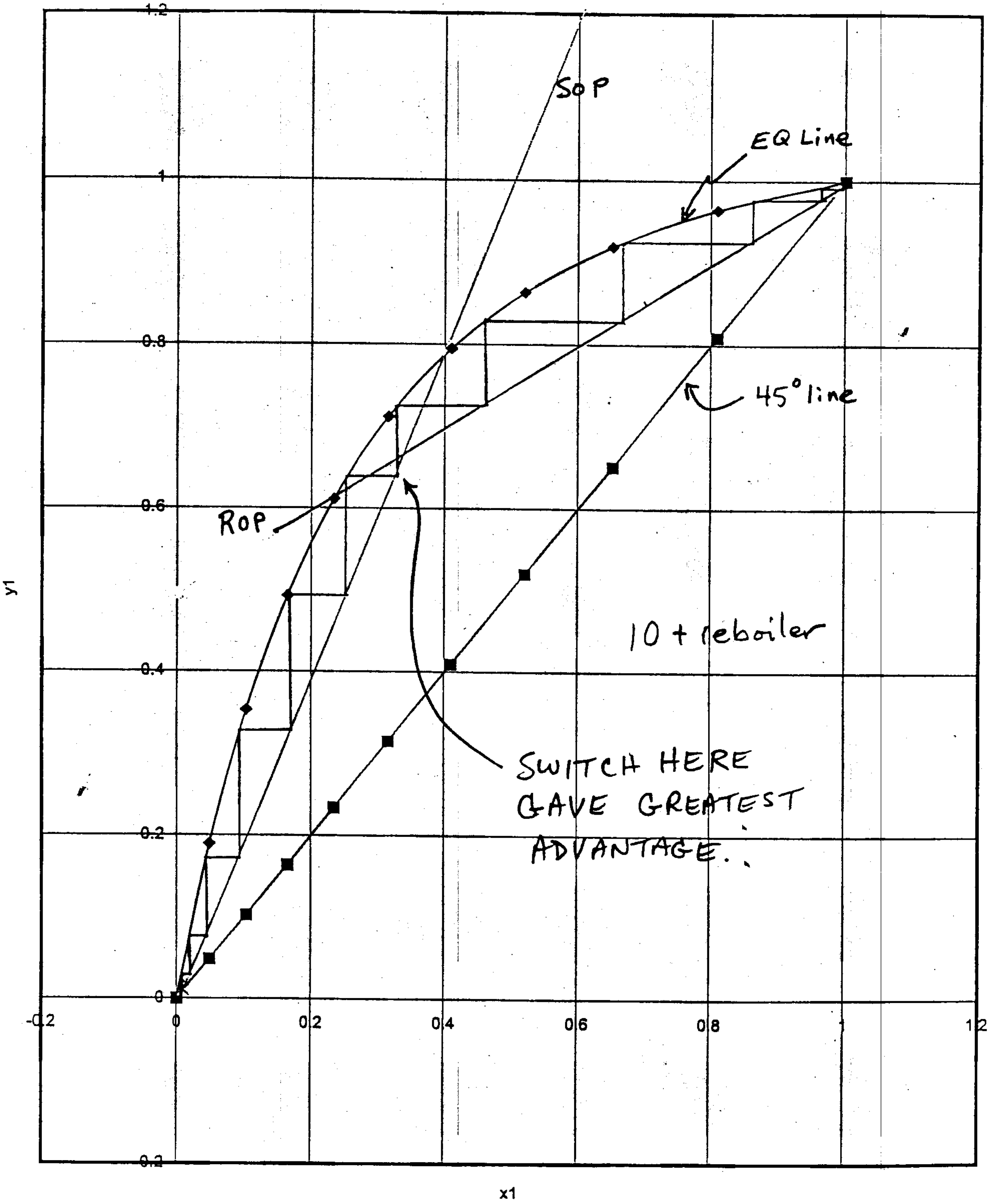
WHY?

LOOKING FOR GREATEST ADVANTAGE

XY Phase Diagram for Hexane/Octane

$R=1$

$V_B=1$



SOP

EQ Line

45° line

ROP

10 + reboiler

SWITCH HERE GAVE GREATEST ADVANTAGE..

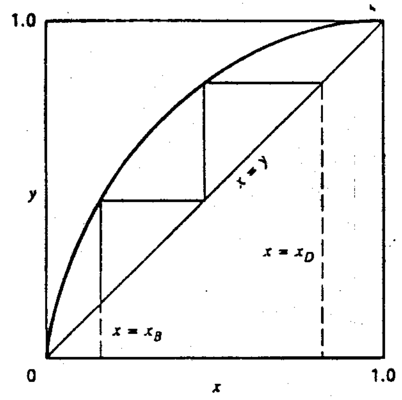
LIMITING CONDITIONS (WHERE PRACTICAL)

① MINIMUM # STAGES

→ WHERE IS GREATEST ADVANTAGE?

∞ REFLUX ⇒ SLOPE = 1 ⇒ 45° LINE

SEADER + HENLEY Fig. 7.11



GREATEST ADVANTAGE BECAUSE ANY REFLUX OR BOILUP WILL CUT OFF THE DIFFERENCE THAT EACH STAGE MAKES TOWARDS SEPARATION

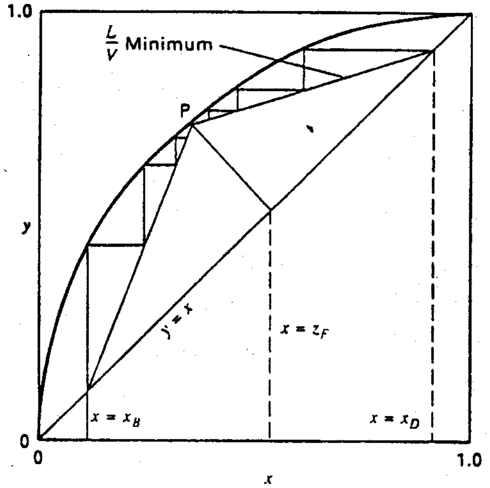
$$\Rightarrow \quad \boxed{N_{MIN} < N < \infty}$$

② MINIMUM REFLUX RATIO

RECTIFIER : WANT $\frac{R}{R+1}$ as Close to 0 as possible (small R)

STRIPPER : WANT $\frac{V_B+1}{V_B}$ as Close to ∞ as possible (small V_B)

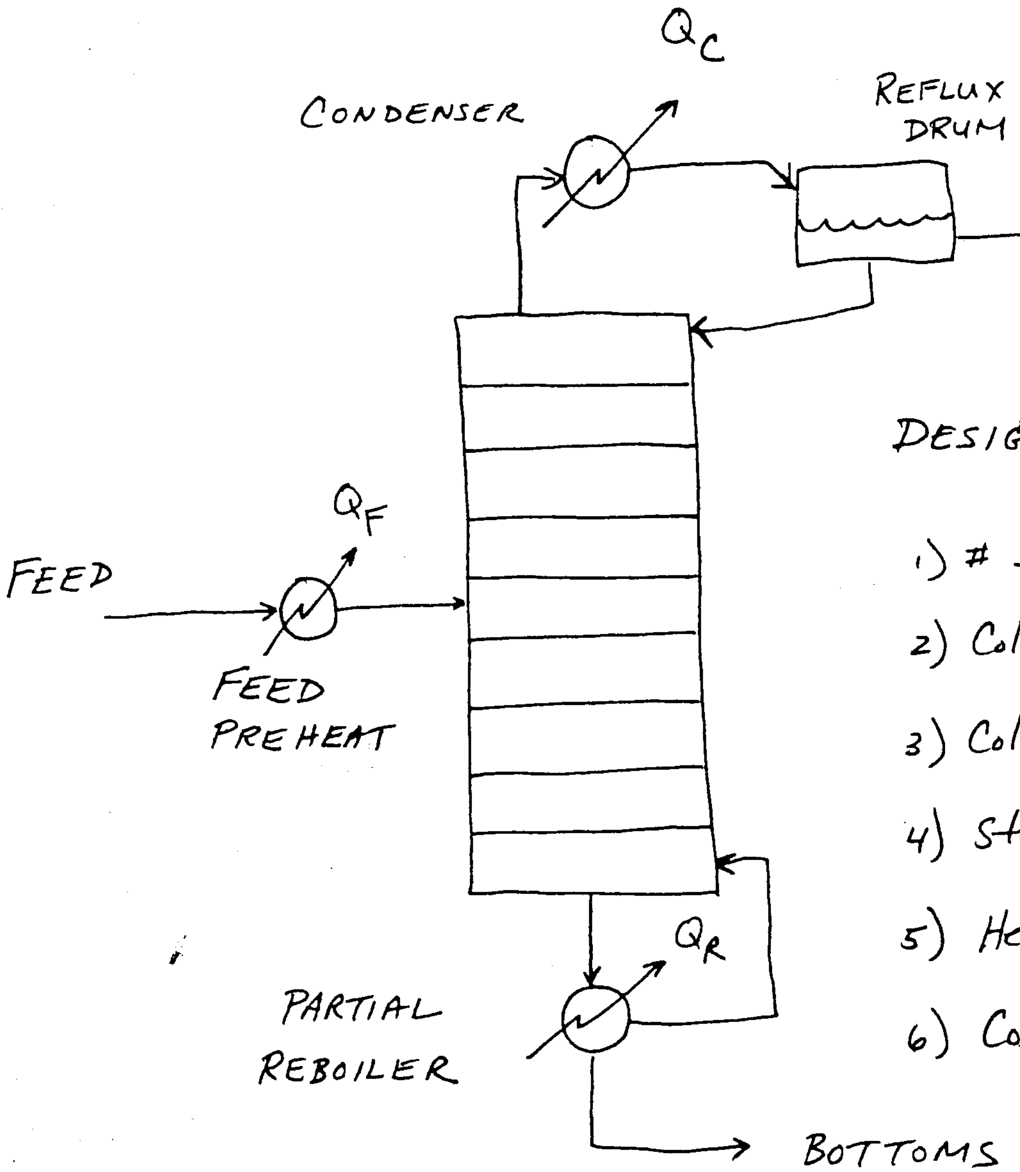
SEADER + HENLEY Fig. 7.12a



SOP, ROP, + q-line intersect on EQ Line

$$\Rightarrow \quad \boxed{R_{MIN} < R < \infty}$$

MORE ON DESIGN OF DISTILLATION:



DESIGN PARAMETERS

1) # stages
2) Column pressure
3) Column pressure drop
4) Stage efficiency
5) Heat duties
6) Column sizing
7) Flooding velocity
8) Weeping
9) Tray Spacing
10) Minimum R
11) Minimum # stages

⑩+⑪ SEE NOTES DATED 3-15-02

① # STAGES

McCABE — THELE ⇒ # theoretical stages (TOTAL)

⇒ $\sum$ (stages in column + partial reboiler + partial condenser),

EXAMPLES: (w/ 10 stages total)

Stages in Column?

a) Stripping section

→ No reboiler

→ Partial reboiler

b), Rectifier

→ No condenser

→ Partial condenser

→ Total condenser

c) Distillation column

→ Partial Reboiler/Total Condenser

In reality...

Need to account for efficiency

(Recall, McCabe-Thiele assumes equilibrium at each stage.)

↳ We'll look at this later.

---

②, ③ Column Pressure + PRESSURE DROP

We assumed $\Delta P = 0$

In reality...

Plate ΔP = 50 - 70 mm H_2O (sieve tray)

$\Rightarrow$ 0.005 bar / plate

$\Rightarrow$ Pretty small

Where generated?

REBOILER

EXAMPLE: 40 plates $\Rightarrow$ 0.2 bar $= \Delta P$

ASSUMPTION NOT BAD.

④ STAGE EFFICIENCY

WHY WOULD McCABE-THIELE BE ~ CORRECT?

- → GOOD CONTACT
- → INCREMENTAL CHANGES
- → SIMPLE BINARY

WHY IS ASSUMPTION OF EQUILIBRIUM NOT CORRECT?

- → LARGE TRAYS (MASS TRANSFER)
- → NOT OPTIMIZED (FLOW RATES)
- → ∞ time required (CLOSE ONLY)

OVERALL EFFICIENCY (η_0)

$$\Rightarrow \text{Actual \# Trays} = \frac{1}{\eta_0}\left(\text{Theoretical \# plates}\right)$$

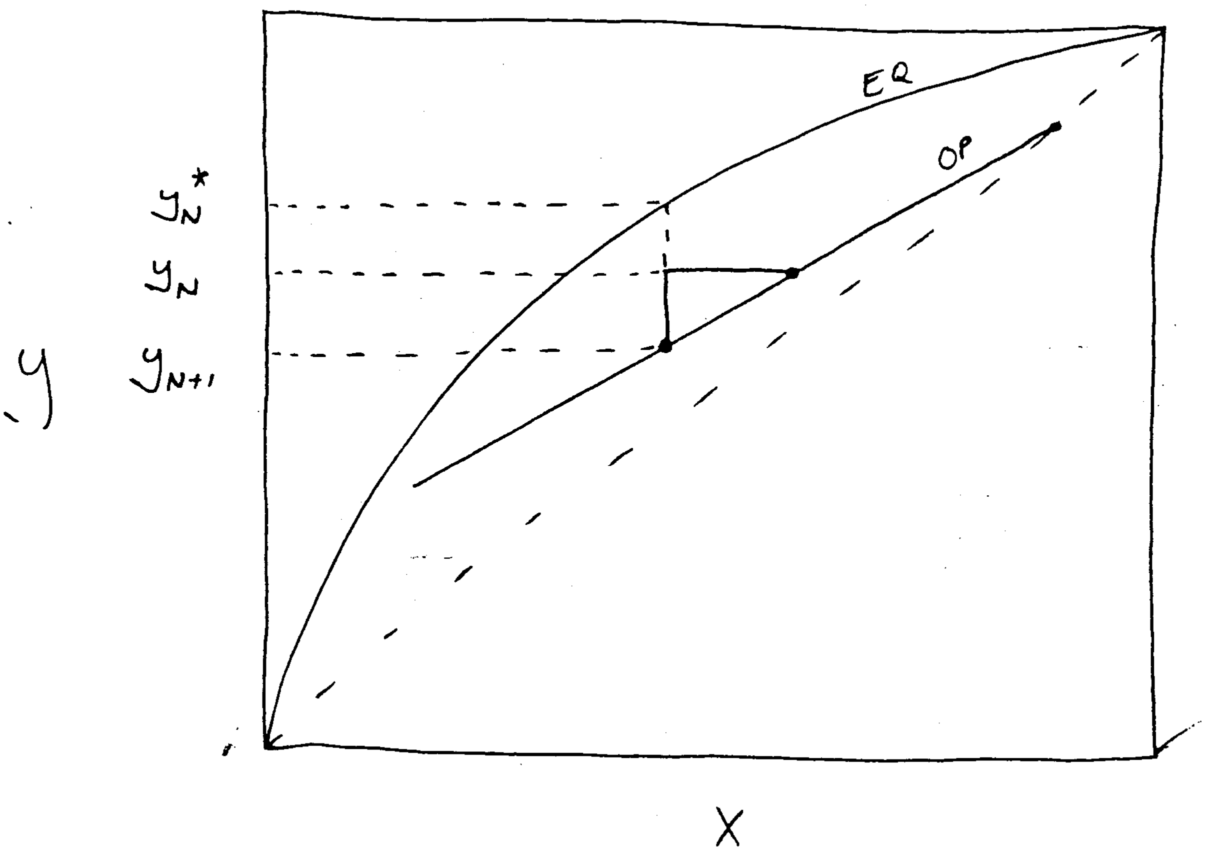
<u>EXAMPLE:</u> $\eta_0 = 60\%$

⇒ McCabe-Thiele shows 6 stages

$$\text{Actual \# Stages} = \frac{6}{0.60} = \underline{\underline{10 \text{ stages}}}$$

MURPHREE EFFICIENCY (η_M)

$$\eta_M = \frac{y_N - y_{N+1}}{y_N^* - y_{N+1}}$$



⇒ IN EITHER CASE, NEED MORE STAGES.

⑤ HEAT DUTIES

TOTAL CONDENSER –

$$Q_C = \underbrace{D(R+1)}_{V \text{ (from mat'l balance)}} \Delta H^{vap}$$

PARTIAL CONDENSER –

$$Q_C = \underbrace{DR}_{L \text{ (from mat'l balance)}} \Delta H^{vap}$$

PARTIAL REBOILER –

$$Q_R = \underbrace{B V_B}_{\bar{V} \text{ (from mat'l balance)}} \Delta H^{vap}$$

<u>NOTE:</u> FLOW RATE TIMES ΔH^{vap}

$\Rightarrow$ Q is rate of heat addition or removal.

UTILITIES:

REBOILER — SATURATED STEAM

$$m_S = \frac{M_S \; Q_R}{\Delta H_S^{vap}}$$

mass flow rate steam used

Molecular weight steam

molar enthalpy of vaporization for steam

CONDENSER — COOLING WATER

$$m_{CW} = \frac{Q_C}{C_{p_{H_2O}} \left(T_{OUT} - T_{IN}\right)}$$

mass flowrate cooling water

Specific heat of water

ΔT for condenser cooling water

RULE OF THUMB:

COST m_S usually 10 X cost m_{CW}

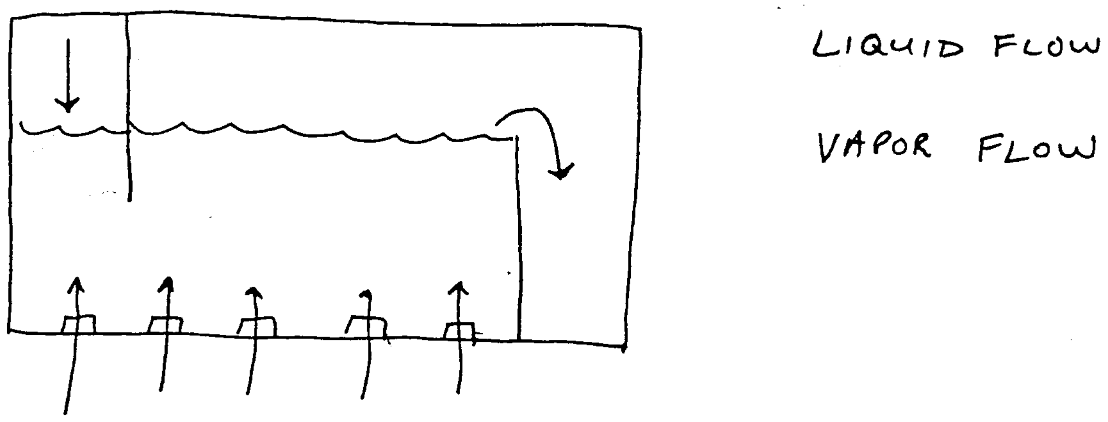
⇒ To help with heat addition, use feed preheat w/ $0 < \psi < 1$.

⑥, ⑦, ⑧, ⑨

COLUMN SIZING, FLOODING VELOCITY, WEEPING, AND TRAY SPACING.

FLOODING VELOCITY –

LINEAR VELOCITY OF THE GAS PHASE.



WHEN GAS FLOW SUFFICIENTLY LARGE, LIQUID CAN NOT FLOW DOWN THE COLUMN ⇒ FLOODING.

Typical to operate close to flooding.
(more efficient contact/mass transfer)

<u>IF</u> GAS VELOCITY TOO LOW ⇒ WEEPING

⇒ Liquid flows through holes in tray.

⇒ <u>Not</u> efficient contact

⇒ Low η_o

$$u_c \propto \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

Flooding velocity

Liquid Density

vapor density

Proportionality Constant, C

$$u_c = C\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

(p. 308 SEADER + HENLEY)

C = f (plant data, plate spacing, flow rates)

Typically rewrite in terms of surface tension, σ

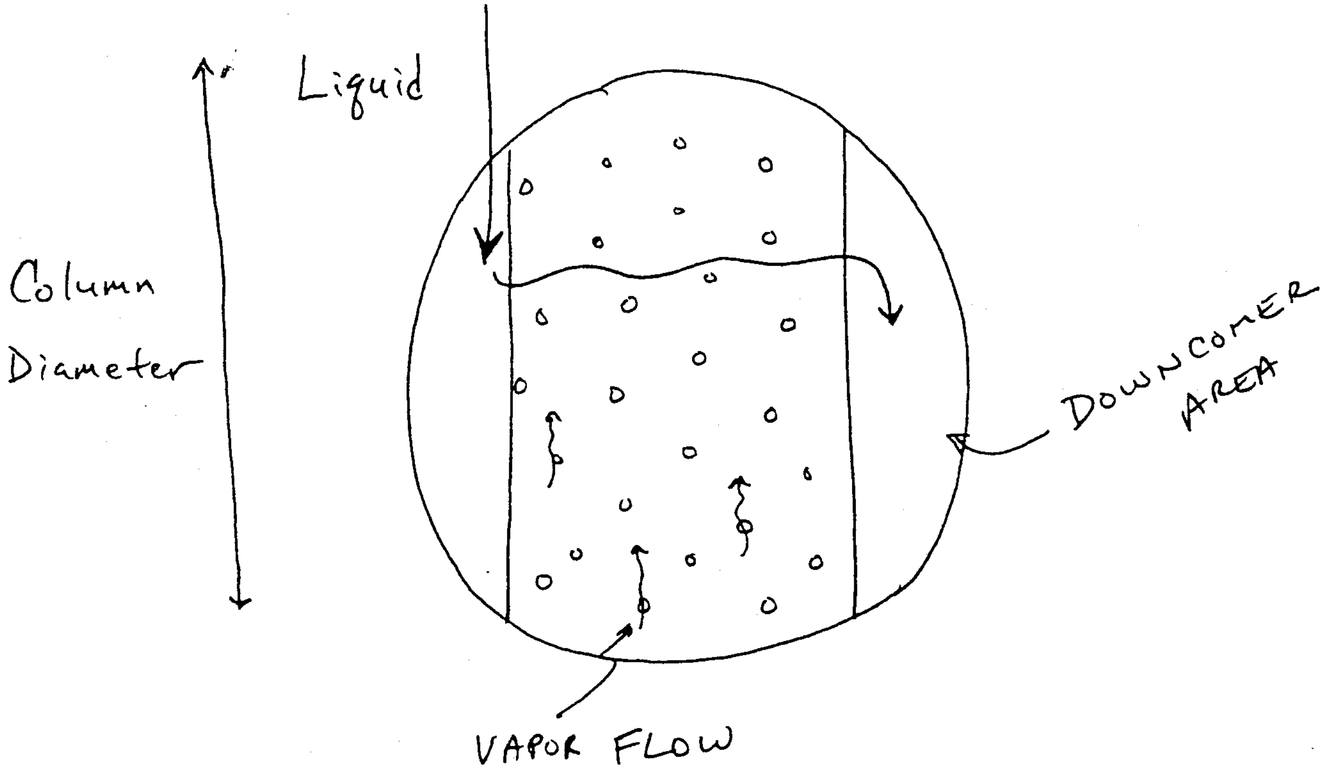
$$u_C = C' \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \left(\frac{\sigma}{20}\right)^{0.2}$$

We'll see this again for Absorption.

ASSUMING WE KNOW u_C (can measure experimentally)

$\Rightarrow$ COLUMN SIZING

LET'S LOOK AT A SIEVE TRAY FROM ABOVE...



DOWN COMER (EACH) AREA: ~ 10 – 15% of area

⇒ 70 – 80% for liquid / vapor contact

WELL, IF WE KNOW V + u_c

1) CONVERT V to Volumetric flow rate

2) $\dfrac{\dot{V}}{u_c}$ = AREA THAT GAS PASSES THROUGH

3) HOWEVER, THIS IS ONLY ~~A~~ THE FRACTION ~~IS~~ USED FOR LIQUID / VAPOR CONTACT. . .

$$\frac{\text{area}}{\text{fraction bubbling}} = \text{total area} = \frac{\pi d^2}{4}$$

4) COLUMN DIAMETER $= \sqrt{\dfrac{4\,(\text{Total Area})}{\pi}}$

Typical: 1 ft – 30 ft

COLUMN HEIGHT:

TRAY SPACING: 150 mm – 2 m
(24" typical)

STAGES; 1 – 100 or so

TRAY SPACING + COLUMN DIAMETER: RELATED

TRAY SPACING ↑ ⇒ COLUMN DIAMETER ↓

+ vice versa

⇒ JUST REMEMBER TO INCLUDE EFFICIENCY WHEN YOU CALCULATE # OF STAGES.

⇒ MASS TRANSFER.