LET'S LOOK AT SOME SIMILARITIES

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

**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**: \( y_m = \left( \frac{\frac{V_B + 1}{V_B}} \right)(x_{m-1}) - \frac{1}{V_B}x_B \)

\[\text{Slope} = \frac{R}{R+1}\]

\[\text{Slope} = \frac{V_B + 1}{V_B}\]
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.
- \( F = 100 \text{ mol/hr} \)
- \( V_B = \frac{V}{B} = 2 \)
- Pressure = 1 bar everywhere.

**STEP 3:** BULK FLOWS

\[
F = \overline{V} + B
\]

\[
100 = \overline{V} + B
\]

(overall mole balance)

\[
V_B = 2 = \frac{V}{B}
\]

\( \Rightarrow \overline{V} = 2B \) (boilup ratio)

\[
100 = 2B + B \Rightarrow B = 33.3 \text{ mol/hr}
\]

\[
\overline{V} = 66.7 \text{ mol/hr}
\]

**STEP 4:** BALANCE AROUND REBOILER

\[
\bar{L} = \overline{V} + B
\]

\( \Rightarrow \bar{L} = 100 \text{ mol/hr} \)
**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)

   \[y_N = 0.46\]
XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar

STRIPPER ONLY

BEST WE CAN DO WITH THIS BOILING RATIO
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 $\Rightarrow$ best we can do
(with given $V_B$)

$\Rightarrow$ Design Change.

BOILUP RATIO

a) $V_B$ smaller

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

b) $V_B$ larger

$$\frac{V_B + 1}{V_B} \Rightarrow 1$$
XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar

- OP Line for $V_B = 0$ 
  ⇒ No stages 
  ⇒ No separation (virtually none)

- OP Line for 
  ⇒ Boilup 
  ⇒ NO PRODUCT (Bottoms)
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.
XY Phase Diagram for a System of Hexane (1) in Octane (2) at 1 bar

DISTILLATION COLUMN
A. FEED CONDITION

POSSIBILITIES

i) Subcooled Liquid (SL)

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).]
RECALL q-line:

\[ \psi = \frac{V}{F} \]  \Rightarrow \text{Flash} \Rightarrow V \text{ is amount (moles) of feed stream vaporized. (NOT VAPOR FLOW RATE IN RECTIFIER)}

Equation q-line:

\[ y = \left( \frac{\psi - 1}{\psi} \right) x + \left( \frac{1}{\psi} \right) z \]

\[ \uparrow \quad \text{Vapor Composition} \quad \text{Liquid Composition} \quad \text{Feed Composition} \]

XY PHASE DIAGRAM EQ line

(Light Key)
FROM PREVIOUS FIGURE...

KNEW

SOP (STRIPPER OPERATING LINE)
ROP (RECTIFIER OPERATING LINE)

INTERSECTION GIVES ONE POINT ON θ-LINE

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

ψ (SLOPE = \(\frac{\psi-1}{\psi}\))

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.

WHY?

LOOKING FOR GREATEST ADVANTAGE
XY Phase Diagram for Hexane/Octane

$R = 1$
$V_B = 1$

10 + reboiler

SWITCH HERE GAVE GREATEST ADVANTAGE.
LIMITING CONDITIONS (WHERE PRACTICAL)

1 MINIMUM + STAGES
   \[ \text{WHERE IS GREATEST ADVANTAGE?} \]
   \[ \infty \text{ REFLUX} \Rightarrow \text{SLOPE} = 1 \Rightarrow 45^\circ \text{LINE} \]

\[ \Rightarrow \quad N_{\text{MIN}} < N < \infty \]

2 MINIMUM REFLUX RATIO
   \[ \text{RECTIFIER: WANT} \quad \frac{R}{R+1} \text{ as close to 0 as possible} \quad \text{(small R)} \]
   \[ \text{STRIPPER: WANT} \quad \frac{V_B+1}{V_B} \text{ as close to } \infty \text{ as possible} \quad \text{(small } V_B \text{)} \]

So,ROP, and q-line intersect \[ \Rightarrow \text{ EQ LINE} \]

\[ \Rightarrow R_{\text{MIN}} < R < \infty \]
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 - Thiele \Rightarrow \# theoretical stages (TOTAL)

\Rightarrow \Sigma \left( \text{stages in column} + \text{partial reboiler} + \text{partial condenser} \right)

EXAMPLES: (w/ 10 stages total)

a) Stripping section
   \Rightarrow \text{No reboiler}
   \Rightarrow \text{Partial reboiler}

b) Rectifier
   \Rightarrow \text{No condenser}
   \Rightarrow \text{Partial condenser}
   \Rightarrow \text{Total condenser}

c) Distillation column
   \Rightarrow \text{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. Column Pressure + Pressure Drop
3. We assumed \( \Delta P = 0 \)

In reality...

Plate \( \Delta P = 50-70 \) mm \( H_2O \) (Sieve tray)

\[ \Rightarrow 0.005 \text{ bar/plate} \]

\[ \Rightarrow \text{Pretty small} \]

Where generated?

Reboiler

\[ \text{Example: 40 plates } \Rightarrow 0.25 \text{ bar } = \Delta P \]

Assumption NOT bad.
STAGE EFFICIENCY

WHY WOULD MCCABE - THIELE BE \text{\it{CORRECT}}?

\rightarrow \text{GOOD CONTACT}

\rightarrow \text{INCREMENTAL CHANGES}

\rightarrow \text{SIMPLE BINARY}

WHY IS ASSUMPTION OF EQUILIBRIUM NOT \text{\it{CORRECT}}?

\rightarrow \text{LARGE TRAYS (MASS TRANSFER)}

\rightarrow \text{NOT OPTIMIZED (FLOW RATES)}

\rightarrow \infty \text{ time required (CLOSE ONLY)}

OVERALL EFFICIENCY \( (\eta_0) \)

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

EXAMPLE: \( \eta_0 = 60\% \)

\[ \Rightarrow \text{McCabe - Thiele shows 6 stages} \]

\[ \text{Actual \#} = \frac{6}{0.60} = 10 \text{ stages} \]
Murphree Efficiency ($\eta_m$)

\[ \eta_m = \frac{y_N - y_{N+1}}{y_N^* - y_{N+1}} \]

=> IN EITHER CASE, NEED MORE STAGES.
5) HEAT DUTIES

**TOTAL CONDENSER** -

\[ Q_c = \frac{D (R + 1) \Delta H_{vap}}{V} \]  

(from mat'1 balance)

**PARTIAL CONDENSER** -

\[ Q_c = \frac{D R \Delta H_{vap}}{L} \]  

(from mat'1 balance)

**PARTIAL REBOILER** -

\[ Q_R = \frac{B V_B \Delta H_{vap}}{\bar{V}} \]  

(from mat'1 balance)

**Note:** Flow RATE TIMES \( \Delta H_{vap} \)

\[ \Rightarrow Q \text{ is rate of heat addition} \]

or removal.
UTILITIES:

REBOILER - Saturated Steam

\[ m_S = \frac{M_S Q_R}{\Delta H_{\text{vap}}} \]

- Mass flow rate of steam
- Molecular weight of steam
- Molar enthalpy of vaporization for steam

CONDENSER - Cooling Water

\[ m_{CW} = \frac{Q_C}{C_p H_2O (T_{out} - T_{in})} \]

- Mass flow rate of cooling water
- Specific heat of water

Rule of Thumb:

Cost \( m_S \) usually \( \frac{10}{1} \) cost \( m_{CW} \)

\[ \Rightarrow \text{To help with heat addition, use feed preheat with } 0 < \psi < 1. \]
COLUMN SIZING, FLOODING VELOCITY, WEEPING, AND TRAY SPACING.

FLOODING VELOCITY -

LINEAR VELOCITY OF THE GAS PHASE.

WHEN GAS FLOW SUFFICIENTLY LARGE, LIQUID CANNOT FLOW DOWN THE COLUMN \( \Rightarrow \) FLOODING.

Typical to operate close to flooding, (more efficient contact/mass transfer)
\[ \text{If Gas Velocity Too Low} \implies \text{Weeping} \]
\[ \implies \text{Liquid flows through holes in tray.} \]
\[ \implies \text{Not efficient contact} \]
\[ \implies \text{Low } \eta \]

\[ U_c \propto \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} \]

Flooding velocity

Proportionality constant, \( C \)

\[ U_c = C \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} \]

(p. 308
SEADER + HENLEY)

\[ C = f (\text{plant data, plate spacing, flow rates}) \]
Typically rewrite in terms of surface tension, \( \sigma \):

\[
\begin{align*}
    u_c &= C' \sqrt{\frac{e_L - e_V}{e_V}} \left( \frac{\sigma}{20} \right)^{0.2} \\
    \text{We'll see this again for Absorption.}
\end{align*}
\]

Assuming we know \( u_c \) (can measure experimentally)

\( \Rightarrow \) Column Sizing

Let's look at a sieve tray from above...
DOWN CORNER AREA: ~ 10-15% of area
\Rightarrow 70-80\% for liquid/vapor contact

WELL, IF WE KNOW V + uc

1) CONVERT V to Volumetric flow rate

\[ \frac{V}{uc} = \text{AREA THAT GAS PASSES THROUGH} \]

2) THIS IS THE FRACTION USED FOR LIQUID / VAPOR CONTACT...

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

4) COLUMN DIAMETER = \sqrt{\frac{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.