Single-Stage Equilibrium Processes

- Degree of Freedom Analysis
- Binary Vapor-Liquid Equilibrium
- Azeotropes in binary systems
- Multicomponent Flash Calculations
- Ternary Liquid-Liquid Systems
Degrees of Freedom: Gibb’s Phase Rule

Example: $C=3$ components in $\mathcal{P}=2$ phases at equilibrium.

\[
\begin{align*}
\text{Variables} & \quad \# \\
\ y & \quad C \\
\ x & \quad C \\
\ T & \quad 1 \\
\ P & \quad 1 \\
2C+2 & = 8
\end{align*}
\]

\[
\begin{align*}
\text{Equations} & \quad \# \\
\ \Sigma & \quad 1 \\
\ \Sigma & \quad 1 \\
\ K & \quad C \\
C+2 & = 5
\end{align*}
\]

\[\mathcal{F} = \mathcal{V} - \varepsilon = 3 \text{ degrees of freedom!}\]

\[\therefore \text{ we must specify 3 variables}\]

Options:
- Set $T, P$ and one mole fraction.
- Set 3 mole fractions and determine $T, P$.
- Set $T$ and 2 mole fractions
- ...

In general:

\[
\begin{align*}
\text{Variables} & \quad \# \\
x & \quad CP \\
T & \quad 1 \\
P & \quad 1 \\
\text{CP+2}
\end{align*}
\]

\[
\begin{align*}
\text{Equations} & \quad \# \\
\ \Sigma & \quad \mathcal{P} \\
K & \quad C(\mathcal{P}-1) \\
\text{CP+2} & \quad C(\mathcal{P}-1)+\mathcal{P}
\end{align*}
\]

Gibbs’ Phase Rule:

\[
\mathcal{F} = C - \mathcal{P} + 2
\]

Note: we may not be able to find a solution if we set an impossible choice of intensive variables.
Examples???
Gibbs Phase Rule & Extensive Variables

Example: single-stage equilibrium with $C$ components in $P=2$ phases

In general, we get $C+P+4$ extra variables and $C+2$ extra equations when we include extensive quantities.

$$\mathcal{F} = \mathcal{V} - \mathcal{E} = C + 4$$

Could set $C-1$ of the $z_i, T_F, P_F, T, P$ and then determine the remaining quantities from the equations.
Binary Vapor-Liquid Equilibrium

Diagram:
- F
- z_A
- T_F
- P_F
- Q
- T, P
- V
- y_A
- L
- x_A
**Tabulated Binary VLE Data**

\[ \mathcal{F} = C - P + 2 \Rightarrow \mathcal{F} = 2 - 2 + 2 = 2 \]

Often data is obtained by fixing \( T \) or \( P \) and \( x_A \) or \( y_A \).

- \( x_A(P, y_A) \) at a given \( T \)
- \( x_A(T, y_A) \) at a given \( P \)

### Water (A) - Glycerol (B) at 1 atm

<table>
<thead>
<tr>
<th>( T ) (C)</th>
<th>( y )</th>
<th>( x )</th>
<th>( \alpha )</th>
</tr>
</thead>
<tbody>
<tr>
<td>100</td>
<td>1</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>104.6</td>
<td>0.9996</td>
<td>0.8846</td>
<td>333</td>
</tr>
<tr>
<td>109.8</td>
<td>0.9991</td>
<td>0.7731</td>
<td>332</td>
</tr>
<tr>
<td>128.8</td>
<td>0.998</td>
<td>0.4742</td>
<td>544</td>
</tr>
<tr>
<td>148.2</td>
<td>0.9964</td>
<td>0.3077</td>
<td>627</td>
</tr>
<tr>
<td>175.2</td>
<td>0.9898</td>
<td>0.1756</td>
<td>456</td>
</tr>
<tr>
<td>207</td>
<td>0.9804</td>
<td>0.0945</td>
<td>481</td>
</tr>
<tr>
<td>244.5</td>
<td>0.8308</td>
<td>0.0491</td>
<td>275</td>
</tr>
<tr>
<td>282.5</td>
<td>0.8308</td>
<td>0.025</td>
<td>191</td>
</tr>
<tr>
<td>290</td>
<td>0</td>
<td>0</td>
<td>-</td>
</tr>
</tbody>
</table>

### Para-xylene (A) - Meta-xylene (B) at 1 atm

<table>
<thead>
<tr>
<th>( T ) (C)</th>
<th>( y )</th>
<th>( x )</th>
<th>( \alpha )</th>
</tr>
</thead>
<tbody>
<tr>
<td>133.335</td>
<td>1</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>138.491</td>
<td>0.8033</td>
<td>0.8</td>
<td>1.0041</td>
</tr>
<tr>
<td>138.644</td>
<td>0.6049</td>
<td>0.6</td>
<td>1.0082</td>
</tr>
<tr>
<td>138.795</td>
<td>0.4049</td>
<td>0.4</td>
<td>1.0123</td>
</tr>
<tr>
<td>138.943</td>
<td>0.2032</td>
<td>0.2</td>
<td>1.016</td>
</tr>
<tr>
<td>139.088</td>
<td>0</td>
<td>0</td>
<td>-</td>
</tr>
</tbody>
</table>

**what do these points represent?**
Binary VLE: Methanol/Water

### SHRT Table 4.1b

<table>
<thead>
<tr>
<th>$T$ (°C)</th>
<th>$y$</th>
<th>$x$</th>
<th>$\alpha$</th>
</tr>
</thead>
<tbody>
<tr>
<td>64.5</td>
<td>1</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>66</td>
<td>0.958</td>
<td>0.9</td>
<td>2.53</td>
</tr>
<tr>
<td>69.3</td>
<td>0.87</td>
<td>0.7</td>
<td>2.87</td>
</tr>
<tr>
<td>73.1</td>
<td>0.779</td>
<td>0.5</td>
<td>3.52</td>
</tr>
<tr>
<td>78</td>
<td>0.665</td>
<td>0.3</td>
<td>4.63</td>
</tr>
<tr>
<td>84.4</td>
<td>0.517</td>
<td>0.15</td>
<td>6.07</td>
</tr>
<tr>
<td>89.3</td>
<td>0.365</td>
<td>0.08</td>
<td>6.61</td>
</tr>
<tr>
<td>93.5</td>
<td>0.23</td>
<td>0.04</td>
<td>7.17</td>
</tr>
<tr>
<td>100</td>
<td>0</td>
<td>0</td>
<td>-</td>
</tr>
</tbody>
</table>

This “width” is directly related to the relative volatility, $\alpha$.

### SHR Figure 4.2a

Saturated vapor

Saturated liquid

### SHR Figure 4.2b

This distance is related to the relative volatility, $\alpha$.

### SHR Figure 4.5

$\alpha_{ij} = \frac{K_i}{K_j}$

Mole fraction of component in vapor, $y$

Mole fraction of methanol in vapor

Mole fraction of methanol in liquid

Mole fraction of component 1 in liquid, $x$
Inverse lever-arm rule:

\[
\frac{V}{L} = \frac{DE}{EF}
\]

Given \(x_{\text{hexane}} = 0.3\),

- What is V/L at B-C?
- What is V/L at D-F?
- What is V/L at G?

- What state is G and B?
- What state is H and A?
- What state is E?
Operating Line (q-line)

 species mole balance: \( Fz_A = Vy_A + Lx_A \)

 overall mole balance: \( F = V + L \)

 Eliminate \( L \) and solve for \( y_A \)... 

\[
y_A = \left[ \frac{V/F - 1}{V/F} \right] x_A + \left[ \frac{1}{V/F} \right] z_A \quad \text{“q-line”}
\]

- goes through \( x_A = y_A = z_A \) (45° line)
- \( 0 \leq V/F \leq 1 \)
- Slope is bounded by \(-\infty\) (\( V=0 \)) and \( 0 \) (\( V=F \)).
- \( V=F \) gives us point “A” (always on 45° line).

Given a molar vaporization % \((V/F)\), we can graphically determine the product compositions.

Example: for \( z_{\text{hexane}}=0.6 \) and 60% vaporization, find the \( y_i \).

1. Locate \( z_A \) on the 45° line (point “A”)
2. From \( V/F \), calculate the slope of the \( q \)-line.
3. Follow the \( q \)-line to determine the equilibrium composition (point “B”)

\[ y_A = \left[ \frac{V/F - 1}{V/F} \right] x_A + \left[ \frac{1}{V/F} \right] z_A \]
Azeotropes in Binary Systems
What is an Azeotrope?

At an azeotrope, $K_A = K_B = 1$, $\alpha = 1$.

• No separation possible at this point!
• If you start on one side of the azeotrope, you can only recover at best a pure stream and the azeotrope - not two pure streams!

Typically occur in systems with close boiling point constituents of different chemical type (e.g. ethanol/water).

• zeotropic systems have a monotonic relationship between $x_A$ and $T$. 
Minimum-boiling Azeotropes

Figure 4.6a:
- What can we say about $\gamma_A$ and $\gamma_B$?
- Positive deviation from Raoult’s law: $K_i = P_i^s / P$, ($K_i = \gamma_i P_i^s / P$).

Figure 4.6b
- What can we say about relative volatility across the azeotrope point?
- Note: we can change the pressure to shift the azeotrope a little bit.
- Figure 4.6a (70°C, 123 kPa) shows composition of $x=0.7$.
- Figure 4.6b & 4.6c shows $x=0.72$ (at 66°C & 101 kPa).
- ethanol-water azeotrope disappears at $P < 9.3$ kPa.
Maximum-boiling Azeotrope

Figure 4.7a:
• What can we say about $\gamma_A$ and $\gamma_B$?
• Negative deviation from Raoult’s law: $K_i = P_i^s / P$, ($K_i = \gamma_i P_i^s / P$).

Again, we see a shift in the azeotrope as we change the operating pressure.
• compare 4.7a with 4.7b & 4.7c.
Predicting Azeotrope Formation

Modified Raoult’s law: \( K_i = \gamma_i \frac{P_i}{P} \)

For an azeotrope, \( K_i = 1 \)

\[ \frac{P}{P^s_i} = \gamma_i = \frac{P}{P_i^s} \]

Modified Raoult’s law at an azeotrope

van-Laar: \[
\begin{align*}
\ln \gamma_1 &= \frac{A_{12}}{\left[1 + (x_1 A_{12})/(x_2 A_{21})\right]^2} \\
\ln \gamma_2 &= \frac{A_{21}}{\left[1 + (x_2 A_{21})/(x_1 A_{12})\right]^2}
\end{align*}
\]

Solve these for \( x_1 \) and \( x_2 \).
(Note: \( x_2 = 1 - x_1 \).)

Example: does ethanol (1) and n-hexane (2) form an azeotrope?

\( A_{12} = 2.409, A_{21} = 1.970 \)

Solution:

1. Choose \( P \).
2. Find \( T_{\text{bubble}} \) or \( T_{\text{dew}} \) and corresponding \( x, y \).
3. See if \( K_i = 1 \).

Up next: flash calculations to get \( T_{\text{bubble}} \) or \( T_{\text{dew}} \).
Multicomponent Flash Calculations
Flash Concepts

Two “modes”
- partial vaporization (typically add heat)
- partial condensation (typically remove heat)

“Isothermal flash”
- add/remove heat such that $T = T_F$
- must determine the “duty” of the heat exchanger to maintain $T$

“Adiabatic flash”
- $Q = 0$, $T \neq T_F$
- Temperature drops (vaporization) or rises (condensation).
- $T$ is unknown, so this is more difficult than isothermal flash.
The Rachford-Rice Equation

**Variables:**
- $x$: species mole balances
- $y$: phase equilibrium
- $z$: energy balance
- $F$: total mole balance

**Equations:**

1. $F z_i = L x_i + V y_i$
2. $y_i = K_i x_i$
3. $\sum_{i=1}^{C} y_i = 1$
4. $\sum_{i=1}^{C} x_i = 1$
5. $h_F F + Q = h_V V + h_L L$
6. $F = V + L$

**Auxiliary equations:**
- $K_i(T, P, y_j, x_j)$
- $h_F(T_F, P_F, z_j)$
- $h_V(T, P, y_j)$
- $h_L(T, P, x_j)$

**3C+8 degrees of freedom**

- We typically know $z_i$, $T_F$, and $P_F$.
- 2 “free” variables.

**Combine first two sets of equations**

- Overall mole balance to eliminate $L$

\[
x_i = \frac{F z_i}{L + V K_i} = \frac{F z_i}{F - V + V K_i} = \frac{z_i}{1 + (K_i - 1) \frac{V}{F}}
\]

**From second equation**

\[
y_i = \frac{K_i z_i}{1 + (K_i - 1) \frac{V}{F}}
\]

Still 2C equations - just different form.

Substitute into $\sum x_i$ and $\sum y_i$ and then subtract to get:

\[
\sum_{i=1}^{C} \left( \frac{K_i z_i}{1 + (K_i - 1) \frac{V}{F}} - \frac{z_i}{1 + (K_i - 1) \frac{V}{F}} \right) = 0 \quad \Psi \equiv \frac{V}{F}
\]

\[
\sum_{i=1}^{C} \frac{z_i (K_i - 1)}{1 + (K_i - 1) \Psi} = 0 \quad \text{Rachford-Rice equation}
\]
Now what?

- For **ideal mixtures**, \( K_i(T, p) \). If we know \( T, p \) we can solve the Rachford-Rice equation for \( \Psi \). After that, we can easily get \( x_i, y_i \).

- For **nonideal mixtures**, \( K_i \) also depends on composition!
  - Either go back and solve a bigger system of equations or use successive substitution.
  - Successive substitution is not highly robust, but is usually good enough for these problems.

Typically, convergence is achieved when

\[
\frac{|\Psi^{(k+1)} - \Psi^{(k)}|}{\Psi^{(k)}} < 10^{-4}
\]

If \( T = T_F \) and \( P = P_F \) then we only have one equation to solve for \( \Psi \).

**Ideal Mixtures:**

1. Determine \( K_i(T, p) \) (using Raoults’s law, graphical techniques, etc.)
2. Solve for \( \Psi \).
3. Calculate \( y_i \) & \( x_i \)

**Non-Ideal Mixtures:**

1. Guess a value for \( K_i \) (e.g. from charts or Raoults’s law)
2. Solve for \( \Psi \).
3. Calculate \( y_i \) & \( x_i \)
4. Calculate \( K_i \) using your favorite model.
5. Solve for \( \Psi \).
6. If \( \Psi \) changed, return to step 3.
7. Update \( y_i \) & \( x_i \) using \( \Psi \) calculated in step 6.
Bubble Point & Dew Point

Rachford-Rice equation

\[
\sum_{i=1}^{C} \frac{z_i (K_i - 1)}{1 + (K_i - 1)\Psi} = 0
\]

- At the bubble point, \( V=0, L=F \) \( \Rightarrow \Psi=0 \).

- At the dew point, \( V=F, L=0 \) \( \Rightarrow \Psi=1 \).

Recall \( K_i = K_i(x_i, y_i, T, P) \)

Use Newtons’ method (or another nonlinear equation solver) to solve for bubble/dew point temperature or pressure.

At equilibrium, the vapor is at its dew point and the liquid is at its bubble point.
Example: bubble point calculation

Bubble point: $V=0$, $L=F$, $\Psi=0$.

Solve $\sum_{i=1}^{C} z_i K_i = 1$ to obtain $T$.

Need a model for $K_i(T)$.

<table>
<thead>
<tr>
<th>Component</th>
<th>Feed kmol/h</th>
<th>$z$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$i$-Butane</td>
<td>8.6</td>
<td>0.0319</td>
</tr>
<tr>
<td>$n$-Butane</td>
<td>215.8</td>
<td>0.7992</td>
</tr>
<tr>
<td>$i$-Pentane</td>
<td>28.1</td>
<td>0.1041</td>
</tr>
<tr>
<td>$n$-Pentane</td>
<td>17.5</td>
<td>0.0648</td>
</tr>
<tr>
<td></td>
<td>270</td>
<td></td>
</tr>
</tbody>
</table>

Empirical Correlation (dePriester Chart):

$$\ln K = \frac{a_1}{T^2} + \frac{a_2}{T} + a_3 + b_1 \ln p + \frac{b_2}{p^2} + \frac{b_3}{p}$$

**NOTE**: $T$ in °R and $p$ in psia!

<table>
<thead>
<tr>
<th>Compound</th>
<th>$a$</th>
<th>$a$</th>
<th>$a$</th>
<th>$b$</th>
<th>$b$</th>
<th>$b$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Isobutane</td>
<td>-1,166,846</td>
<td>0</td>
<td>7.72668</td>
<td>-0.92213</td>
<td>0</td>
<td>0</td>
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<tr>
<td>$n$-Butane</td>
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<td>7.94986</td>
<td>-0.96455</td>
<td>0</td>
<td>0</td>
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<tr>
<td>Isopentane</td>
<td>-1,481,583</td>
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<td>7.58071</td>
<td>-0.93159</td>
<td>0</td>
<td>0</td>
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<tr>
<td>$n$-Pentane</td>
<td>-1,524,891</td>
<td>0</td>
<td>7.33129</td>
<td>-0.89143</td>
<td>0</td>
<td>0</td>
</tr>
</tbody>
</table>

M. L. McWilliams, Chemical Engineering, 80(25), 1973 p. 138.
Using Raoult’s law, obtain a $T$-$x$-$y$ diagram of the Propane-Benzene system.

**Raoult’s law:** \[ K_i = \frac{P_i^s}{P} \]

**Modified Raoult’s law:** \[ K_i = \gamma_i \frac{P_i^s}{P} \]

**Antoine Equation parameters**

<table>
<thead>
<tr>
<th></th>
<th>A</th>
<th>B</th>
<th>C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Propane</td>
<td>6.80398</td>
<td>803.81</td>
<td>246.99</td>
</tr>
<tr>
<td>Benzene</td>
<td>6.90565</td>
<td>1211.033</td>
<td>220.79</td>
</tr>
</tbody>
</table>
Isothermal Flash

Given: $T=T_F$, $P=P_F$,

Find: $x_i, y_i, V, L, Q$.

Once we know $\Psi$,

$$x_i = \frac{z_i}{1 + (K_i - 1)\Psi}$$

$$y_i = K_i x_i = \frac{K_i z_i}{1 + (K_i - 1)\Psi}$$

$$V = \Psi F$$

$$L = F - V = F(1 - \Psi)$$

$$Q = h_V V + h_L L - h_F F$$

$$h_i = h_i(T)$$

$$h = \sum_{i=1}^{C} h_i x_i$$ (molar enthalpy)

1. Choose a model for $K_i$.
2. Given $z_i$, solve Rachford-Rice to obtain $\Psi$.
3. Calculate $x_i, y_i$, $V = \Psi F$, $L = F(1 - \Psi)$.
4. Calculate $h_V, h_L, h_F$ at $T$ and $x_i, y_i$
5. Calculate $Q$ (heat exchanger duty).
Adiabatic Flash

Given: $Q=0, P=P_F$

Find: $x_i, y_i, T, V, L.$

**Energy balance:**
$$Q = h_V V + h_L L - h_F F$$
$$0 = F (h_V \Psi + h_L (1 - \Psi) - h_F)$$

**Rachford-Rice equation:**
$$\sum_{i=1}^{C} \frac{z_i (K_i - 1)}{1 + (K_i - 1) \Psi} = 0$$

2 equations, 2 unknowns ($\Psi, T$)

**Energy Balance:**
$$F [h_V \Psi + h_L (1 - \Psi) - h_F] = 0$$

**Note:** at each iteration, we must update values for $h_V$ and $h_L$ given the current value for $T$.
**Summary - Flash Calculations**

<table>
<thead>
<tr>
<th>Description</th>
<th>Knowns</th>
<th>Unknowns</th>
<th>Problem Statement</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bubble point pressure</td>
<td>$T, z_i$</td>
<td>$P$</td>
<td>Solve $r(P) = 1 - \sum_{i=1}^{C} z_i K_i$ for $P$</td>
</tr>
<tr>
<td>Bubble point temperature</td>
<td>$P, z_i$</td>
<td>$T$</td>
<td>Solve $r(T) = 1 - \sum_{i=1}^{C} z_i K_i$ for $T$</td>
</tr>
<tr>
<td>Dew point pressure</td>
<td>$T, z_i$</td>
<td>$P$</td>
<td>Solve $r(P) = 1 - \sum_{i=1}^{C} z_i K_i / K_i$ for $P$</td>
</tr>
<tr>
<td>Dew point temperature</td>
<td>$P, z_i$</td>
<td>$T$</td>
<td>Solve $r(T) = 1 - \sum_{i=1}^{C} z_i K_i / K_i$ for $T$</td>
</tr>
<tr>
<td>Isothermal flash</td>
<td>$T, P, z_i$</td>
<td>$\Psi, x_i, y_i$</td>
<td>Solve the Rachford-Rice equation $r(\Psi) = \sum_{i=1}^{C} \frac{z_i (K_i - 1)}{1 + (K_i - 1) \Psi}$ together with $x_i = \frac{z_i}{1 + (K_i - 1) \Psi}$ and $y_i = K_i x_i$.</td>
</tr>
<tr>
<td>Adiabatic flash</td>
<td>$Q, P, z_i$</td>
<td>$\Psi, T, x_i, y_i$</td>
<td>Solve the coupled system of equations $r_1(\Psi, T) = \sum_{i=1}^{C} \frac{z_i (K_i - 1)}{1 + (K_i - 1) \Psi}$, $r_2(\Psi, T) = F [h_V \Psi + h_L (1 - \Psi) - h_F]$ together with $x_i = \frac{z_i}{1 + (K_i - 1) \Psi}$ and $y_i = K_i x_i$.</td>
</tr>
</tbody>
</table>
Ternary Liquid-Liquid Systems

Graphical Methods
We have a carrier (A) which holds solute (B). We want to extract the solute into a solvent (C).

“**A**” is insoluble in solvent “**C**”

“**simple**” mass balance.

“**A**” is partly soluble in solvent “**C**”

need to know phase equilibrium.
Case 1: Carrier (A) is insoluble in solvent (C)

Carrier (A) mole balance:
\[ x_A^{(F)} F = x_A^{(R)} R \]

Solute (B) mole balance:
\[ x_B^{(F)} F = x_B^{(E)} E + x_B^{(R)} R \]

Solvent (C) mole balance:
\[ S = x_C^{(E)} E \]

Use A & C balances to eliminate R & E in B balance:
\[ x_B^{(F)} F = x_B^{(E)} \frac{S}{x_C} + x_B^{(R)} \frac{F_A}{x_A} \]
\[ X_B^{(F)} F_A = X_B^{(E)} S + X_B^{(R)} F_A \]

Molar ratios of B to “other” component:
\[ X_B^{(E)} = \frac{x_B^{(E)} E}{x_C} = \frac{x_B^{(E)}}{x_C} \text{ in extract, only B & C} \]
\[ X_B^{(R)} = \frac{x_B^{(R)} R}{x_A} = \frac{x_B^{(R)}}{x_A} \text{ in raffinate, only B & A} \]

Distribution or partition coefficient
\[ K'_{DB} = \frac{X_B^{(E)}}{X_B^{(R)}} = \frac{x_B^{(E)}/x_C}{x_B^{(R)}/x_A} = K_{DB} \frac{1 - x_B^{(R)}}{1 - x_B^{(E)}} \]

This is the familiar “K-value” (distribution coefficient) for liquid-liquid.

\[ K_{DB} = \frac{x_B^{(E)}}{x_B^{(R)}} \]

Extraction factor (large is good)
\[ \frac{X_B^{(R)}}{X_B^{(F)}} = \left(1 + \frac{K'_{DB} S}{F_A}\right)^{-1} \]
Case 2: Carrier (A) is soluble in solvent (C)

1. Locate feed (M)
2. Follow tie-lines to miscibility boundary to determine product composition (E & R)
3. Inverse lever-arm rule to determine relative amount of E & R.

Identify/describe:
- How many degrees of freedom?
- Miscibility boundary
- Two-phase region
- Single-phase region
- Tie line
- Plait point
- How is this diagram made?

Note: these diagrams are for a specific $T, P$. 
Example

Determine the extract and raffinate compositions when a 45 wt% glycol, 55 wt% water solution is contacted with twice its weight of pure furfural solvent at 25 °C and 101 kPa.

Assume $F = 100$ g basis of feed

- $S = 200$ g.
- Point “F” indicates F stream
- Point “S” indicates S stream
- Inverse lever arm rule to get point M: $(SM)/(MF) = 1/2$ or $(SM)/(SF) = 1/3$
- Follow tie line to get E & R.
  - Extract: 4% A, 9% B, 88% C
  - Raffinate: 56% A, 34% B, 10% C
- Inverse lever arm rule on tie line to get ratio of E/M and R/M:
  - $E = M (MR)/(ER) = 300 \times (5.5 \text{ cm} / 7.6 \text{ cm}) = 220$ g (measured from SHR figure 4.14)
  - $R = M (EM)/(ER) = M-E = 300 - 220 = 80$ g.
Variations on the Theme

“Right-triangle” diagram
- 2 of the 3 components
- same information as equilateral triangle diagram
- often simpler to deal with.

“Equilibrium solute” diagram
- 2 of the 3 components
- plots tie-line endpoint information.