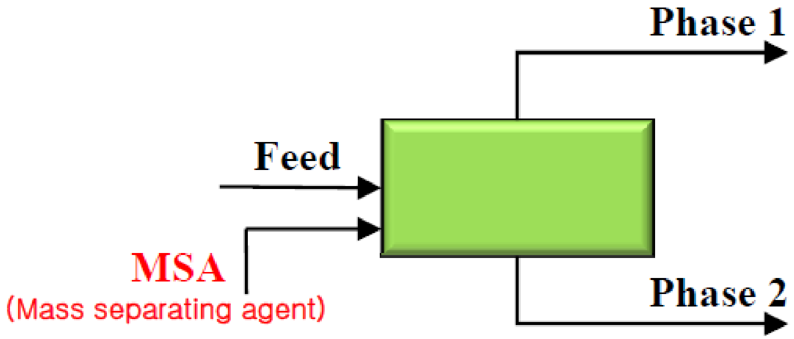
Extraction

Principal references: Chapter 12 in C.J. Geankoplis book and Chapter 8 in Henley, Seader & Roper book

Overview and definitions



- Liquid-Liquid Extraction (LLE)
  - (solvent extraction)

Based on phase-addition

- Pioneered during 1940’s (uranium purification)
- Alternative to distillation, absorption/stripping
  - Energy savings
  - Sometimes easier separation
  - Lower temperatures
- Usually two distinct phases formed:
  - light phase
  - heavy phase

Overview and definitions

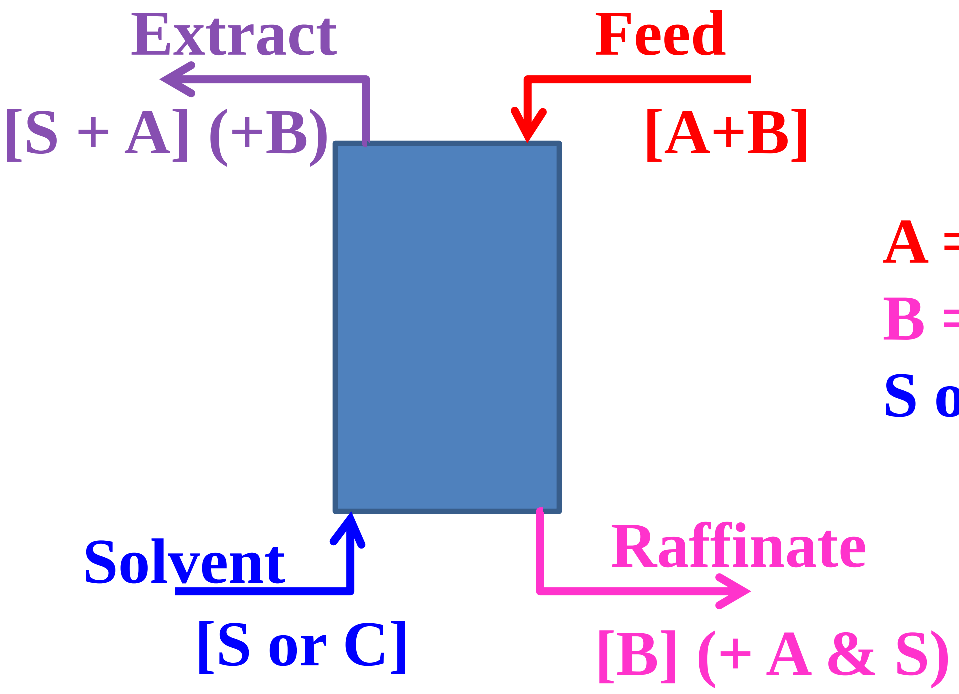
Where LLE is used?

- Bioseparations.
- Nuclear (uranium recovery).
- Mining: nickel/cobalt; copper/iron.
- Perfumes, fragrances and essential oils.
- Fine and specialty chemicals.
- Oil sands: extract bitumen away from water.

Why LLE?

- Temperature sensitive products.
- High purity requirements.
- High-boiling point species in low quantity.
- Need to separate by species type (rather than relative volatility).Close-boiling points, but high solubility difference.
- Azeotrope-forming mixtures.

Overview and definitions



Extract

[S + A] (+B)

Feed

[A+B]

Solvent

[S or C]

Raffinate

[B] (+ A & S)

A = solute

B = Carrier

S or C = solvent

let: $x \equiv$ mass fraction solute in raffinate phase

$y \equiv$ mass fraction solute in extract phase

$y^* =$ in equilibrium with associated x

Main objectives of LLE:

- High recovery of solute overall (low x_R and high y_E).
- Concentrated solute in extract (high y_E)

Overview and definitions

- **Solute:** species to be recovered from the feed
- **Feed:** “feed solvent": “carrier“ of solute.
- **Solvent**: MSA (by convention: the “added" liquid).
- **Extract:** solvent (not solute) mostly present in this layer.

$y_{E,A}$ or y_A = concentration of A, the solute, in extract.

We aim for the solute (A) to be mostly in the extract stream.

Raffinate: residual solute in this layer.

$x_{R,A}$ or x_A= concentration of A, the solute, in raffinite.

Distribution coefficient: quantitative measure of how the solute partitions itself at equilibrium:

$$\boxed{K_A = y_A / x_A}$$

- Measure affinity of solute
- From thermodynamic view:

$$\boxed{K_A = (\mu_A^R - \mu_A^E)/(RT)}$$

Overview and definitions

- **Relative selectivity, β**, of solute with respect to the carrier:

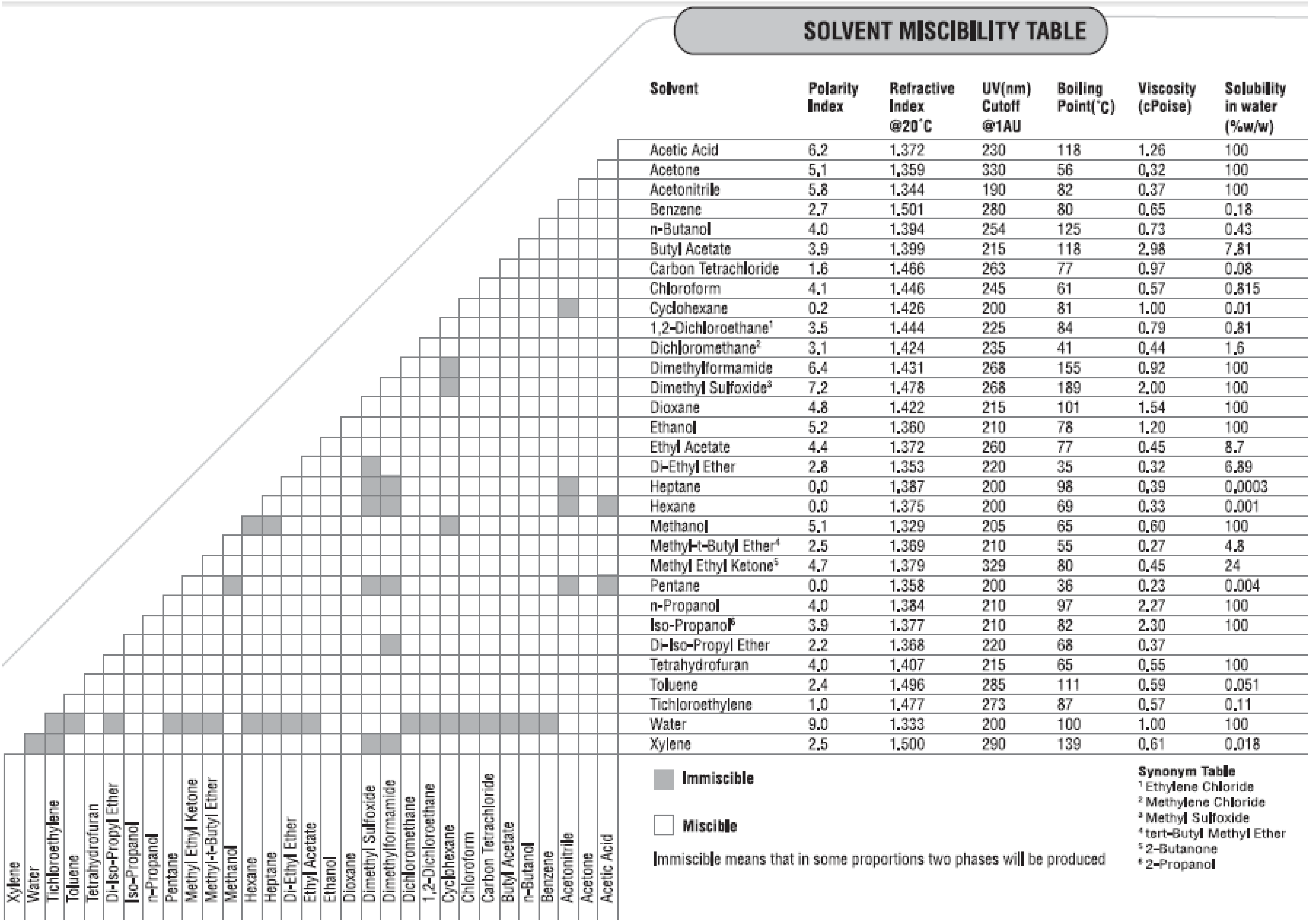
$$\boxed{\beta = \frac{K_A}{K_B} = \frac{y_A / x_A}{y_B / x_B}}$$

- Extraction unit (extractor) could be:
  - Column in stages or packing.
  - Column with moving internals.
  - single stage mixer/settler.
  - Equilibrium stage(s).

- LLE is usually coupled with another separation technique.

- Solvent miscibility is one of the main factors that play key role in LLE process.

Topic II. Extraction



SOLVENT MISCIBILITY TABLE

Solvent	Polarity Index	Refractive Index @20°C	UV(nm) Cutoff @1AU	Boiling Point(°C)	Viscosity (cPoise)	Solubility in water (%w/w)
Acetic Acid	6.2	1.372	230	118	1.26	100
Acetone	5.1	1.359	330	56	0.32	100
Acetonitrile	5.8	1.344	190	82	0.37	100
Benzene	2.7	1.501	280	80	0.65	0.18
n-Butanol	4.0	1.394	254	125	0.73	0.43
Butyl Acetate	3.9	1.399	215	118	2.98	7.81
Carbon Tetrachloride	1.6	1.466	263	77	0.97	0.08
Chloroform	4.1	1.446	245	61	0.57	0.815
Cyclohexane	0.2	1.426	200	81	1.00	0.01
1,2-Dichloroethane[1]	3.5	1.444	225	84	0.79	0.81
Dichloromethane[2]	3.1	1.424	235	41	0.44	1.6
Dimethylformamide	6.4	1.431	268	155	0.92	100
Dimethyl Sulfoxide[3]	7.2	1.478	268	189	2.00	100
Dioxane	4.8	1.422	215	101	1.54	100
Ethanol	5.2	1.360	210	78	1.20	100
Ethyl Acetate	4.4	1.372	260	77	0.45	8.7
Di-Ethyl Ether	2.8	1.353	220	35	0.32	6.89
Heptane	0.0	1.387	200	98	0.39	0.0003
Hexane	0.0	1.375	200	69	0.33	0.001
Methanol	5.1	1.329	205	65	0.60	100
Methyl-t-Butyl Ether[4]	2.5	1.369	210	55	0.27	4.8
Methyl Ethyl Ketone[5]	4.7	1.379	329	80	0.45	24
Pentane	0.0	1.358	200	36	0.23	0.004
n-Propanol	4.0	1.384	210	97	2.27	100
Iso-Propanol[6]	3.9	1.377	210	82	2.30	100
Di-Iso-Propyl Ether	2.2	1.368	220	68	0.37	
Tetrahydrofuran	4.0	1.407	215	65	0.55	100
Toluene	2.4	1.496	285	111	0.59	0.051
Tichloroethylene	1.0	1.477	273	87	0.57	0.11
Water	9.0	1.333	200	100	1.00	100
Xylene	2.5	1.500	290	139	0.61	0.018

Solvent	Xylene	Water	Tichloroethylene	Toluene	Tetrahydrofuran	Di-Iso-Propyl Ether	Iso-Propanol	n-Propanol	Pentane	Methyl Ethyl Ketone	Methyl-t-Butyl Ether	Methanol	Hexane	Heptane	Di-Ethyl Ether	Ethyl Acetate	Ethanol	Dioxane	Dimethyl Sulfoxide	Dimethylformamide	Dichloromethane	1,2-Dichloroethane	Cyclohexane	Chloroform	Carbon Tetrachloride	Butyl Acetate	n-Butanol	Benzene	Acetonitrile	Acetone	Acetic Acid
Acetic Acid																															
Acetone																															☐
Acetonitrile																														☐	☐
Benzene																													☐	☐	☐
n-Butanol																												☐	☐	☐	☐
Butyl Acetate																											☐	☐	☐	☐	☐
Carbon Tetrachloride																										☐	☐	☐	☐	☐	☐
Chloroform																									☐	☐	☐	☐	☐	☐	☐
Cyclohexane																								☐	☐	☐	☐	☐	▨	☐	☐
1,2-Dichloroethane[1]																							☐	☐	☐	☐	☐	☐	☐	☐	☐
Dichloromethane[2]																						☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Dimethylformamide																					☐	☐	▨	☐	☐	☐	☐	☐	☐	☐	☐
Dimethyl Sulfoxide[3]																				☐	☐	☐	▨	☐	☐	☐	☐	☐	☐	☐	☐
Dioxane																			☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Ethanol																		☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Ethyl Acetate																	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Di-Ethyl Ether																☐	☐	☐	▨	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Heptane															☐	☐	☐	☐	▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	▨	☐	☐
Hexane														☐	☐	☐	☐	☐	▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	▨	☐	▨
Methanol													▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	▨	☐	☐	☐	☐	☐	☐	☐	☐
Methyl-t-Butyl Ether[4]												☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Methyl Ethyl Ketone[5]											☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Pentane										☐	☐	▨	☐	☐	☐	☐	☐	☐	▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	▨	☐	▨
n-Propanol									☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Iso-Propanol[6]								☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Di-Iso-Propyl Ether							☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	▨	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Tetrahydrofuran						☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Toluene					☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Tichloroethylene				☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐
Water			▨	▨	☐	▨	☐	☐	▨	▨	▨	☐	▨	▨	▨	▨	☐	☐	☐	☐	▨	▨	▨	▨	▨	▨	▨	▨	☐	☐	☐
Xylene		▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	▨	▨	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐	☐

▨ **Immiscible**

☐ **Miscible**

Immiscible means that in some proportions two phases will be produced

Synonym Table

[1] Ethylene Chloride
[2] Methylene Chloride
[3] Methyl Sulfoxide
[4] tert-Butyl Methyl Ether
[5] 2-Butanone
[6] 2-Propanol

Overview and definitions

Three major steps required in LLE:

1. Mixing/contacting:

- turbulent contact between liquid phases
- small droplet dispersion in a continuous phase
- which phase is dispersed?
- mass-transfer between phases
- limited by solute loading in solvent

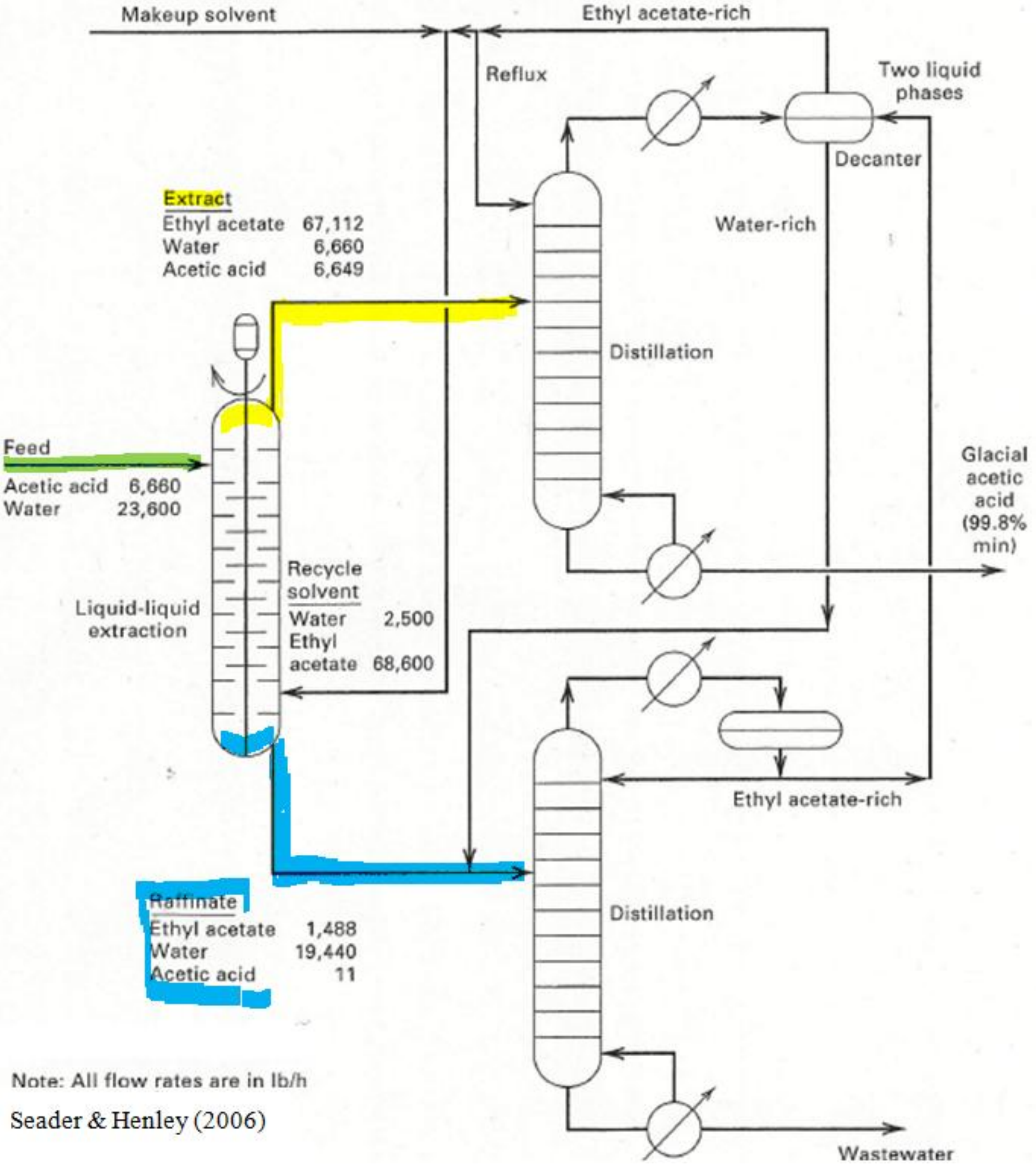
2. Phase separation:

- reverse of above mixing step
- drops come together and coalesce
- relies on density difference

3. Collection of phases leaving the unit

- split the raffinate from the extract

Typical LLE Process



Note: All flow rates are in lb/h

Seader & Henley (2006)

Examples of Industrial LLE Processes

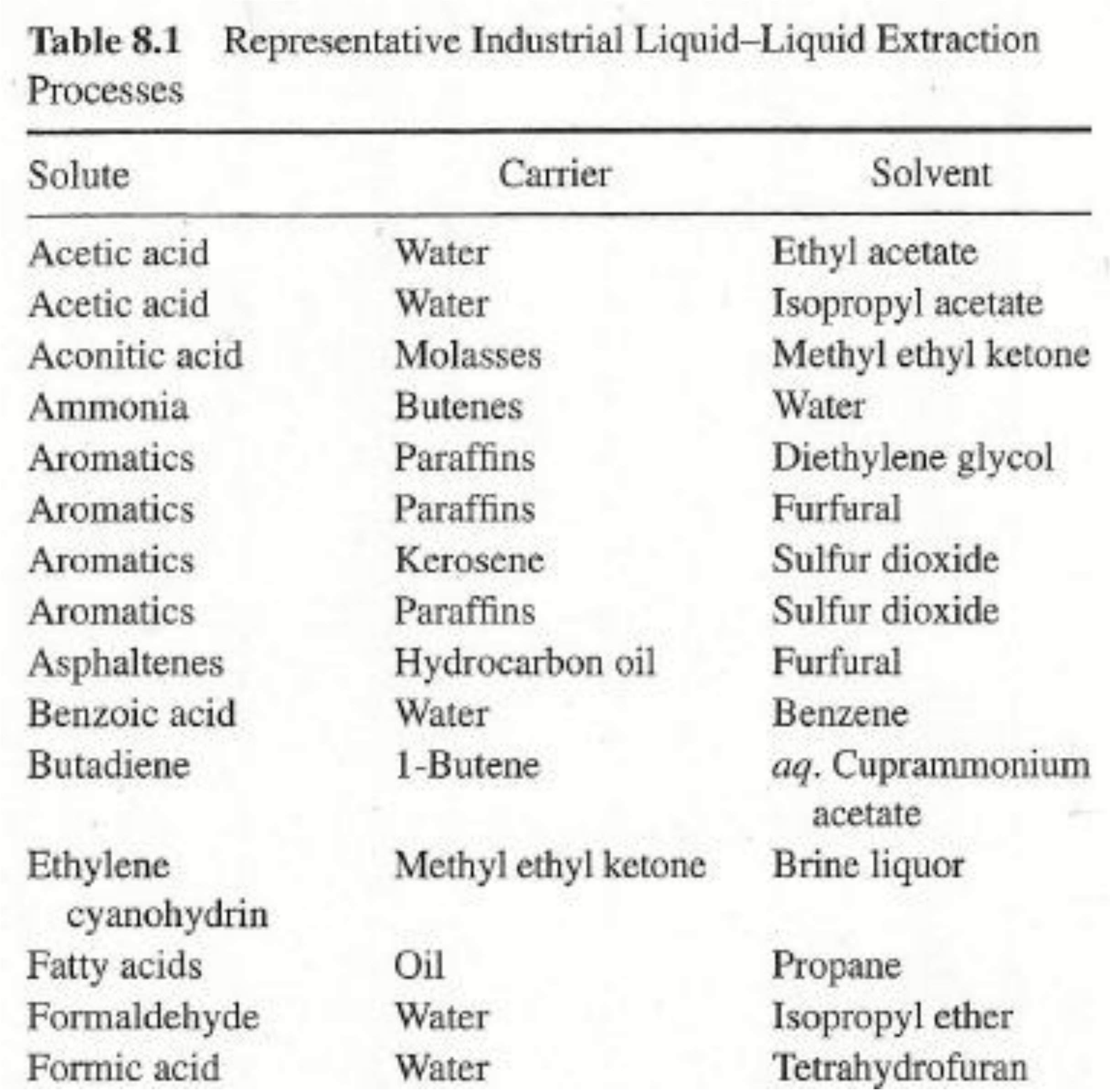
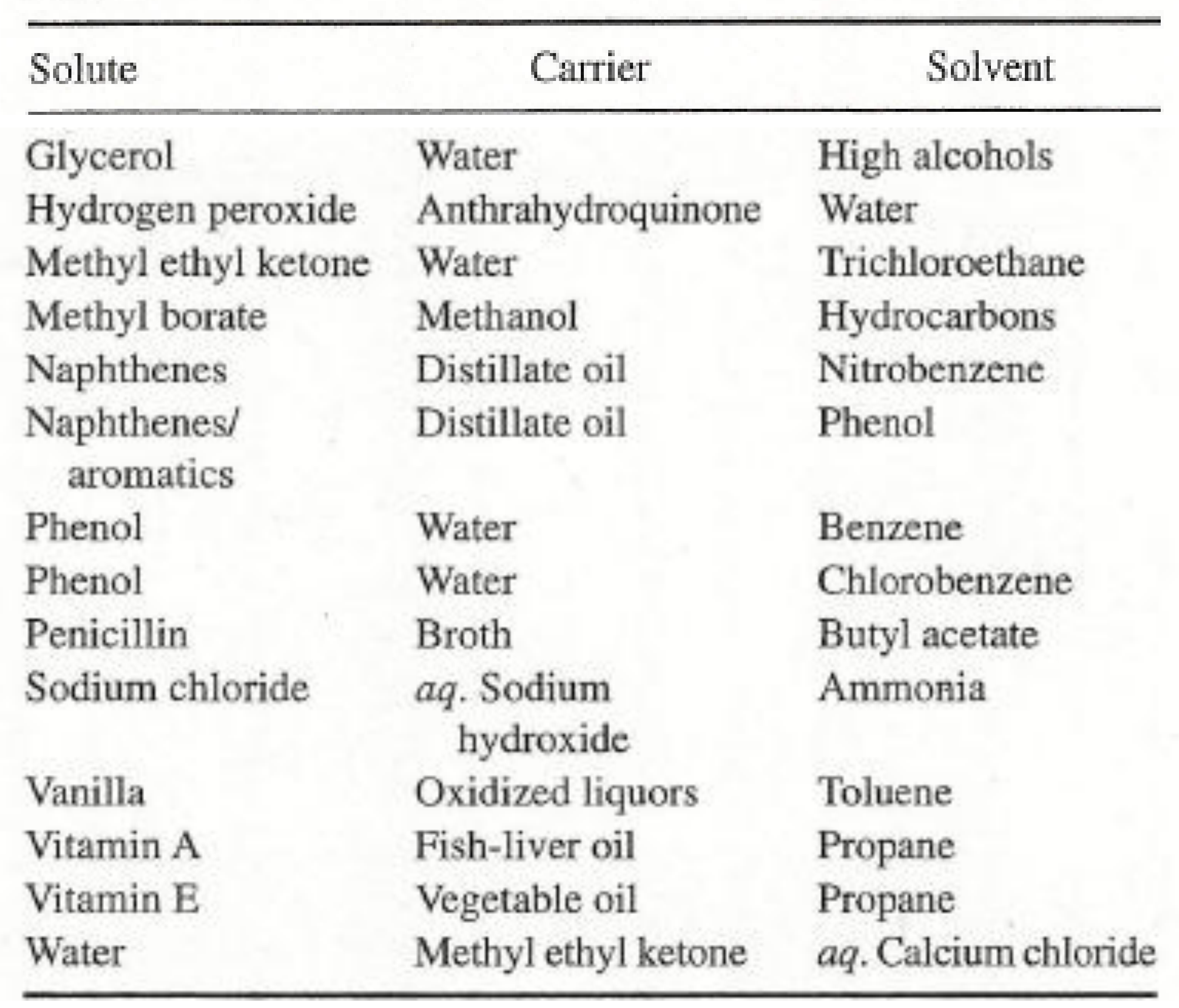


Table 8.1 Representative Industrial Liquid–Liquid Extraction Processes

Solute	Carrier	Solvent
Acetic acid	Water	Ethyl acetate
Acetic acid	Water	Isopropyl acetate
Aconitic acid	Molasses	Methyl ethyl ketone
Ammonia	Butenes	Water
Aromatics	Paraffins	Diethylene glycol
Aromatics	Paraffins	Furfural
Aromatics	Kerosene	Sulfur dioxide
Aromatics	Paraffins	Sulfur dioxide
Asphaltenes	Hydrocarbon oil	Furfural
Benzoic acid	Water	Benzene
Butadiene	1-Butene	*aq.* Cuprammonium acetate
Ethylene cyanohydrin	Methyl ethyl ketone	Brine liquor
Fatty acids	Oil	Propane
Formaldehyde	Water	Isopropyl ether
Formic acid	Water	Tetrahydrofuran

Examples of Industrial LLE Processes



Solute	Carrier	Solvent
Glycerol	Water	High alcohols
Hydrogen peroxide	Anthrahydroquinone	Water
Methyl ethyl ketone	Water	Trichloroethane
Methyl borate	Methanol	Hydrocarbons
Naphthenes	Distillate oil	Nitrobenzene
Naphthenes/aromatics	Distillate oil	Phenol
Phenol	Water	Benzene
Phenol	Water	Chlorobenzene
Penicillin	Broth	Butyl acetate
Sodium chloride	*aq*. Sodium hydroxide	Ammonia
Vanilla	Oxidized liquors	Toluene
Vitamin A	Fish-liver oil	Propane
Vitamin E	Vegetable oil	Propane
Water	Methyl ethyl ketone	*aq*. Calcium chloride

Selecting a solvent for LLE Processes

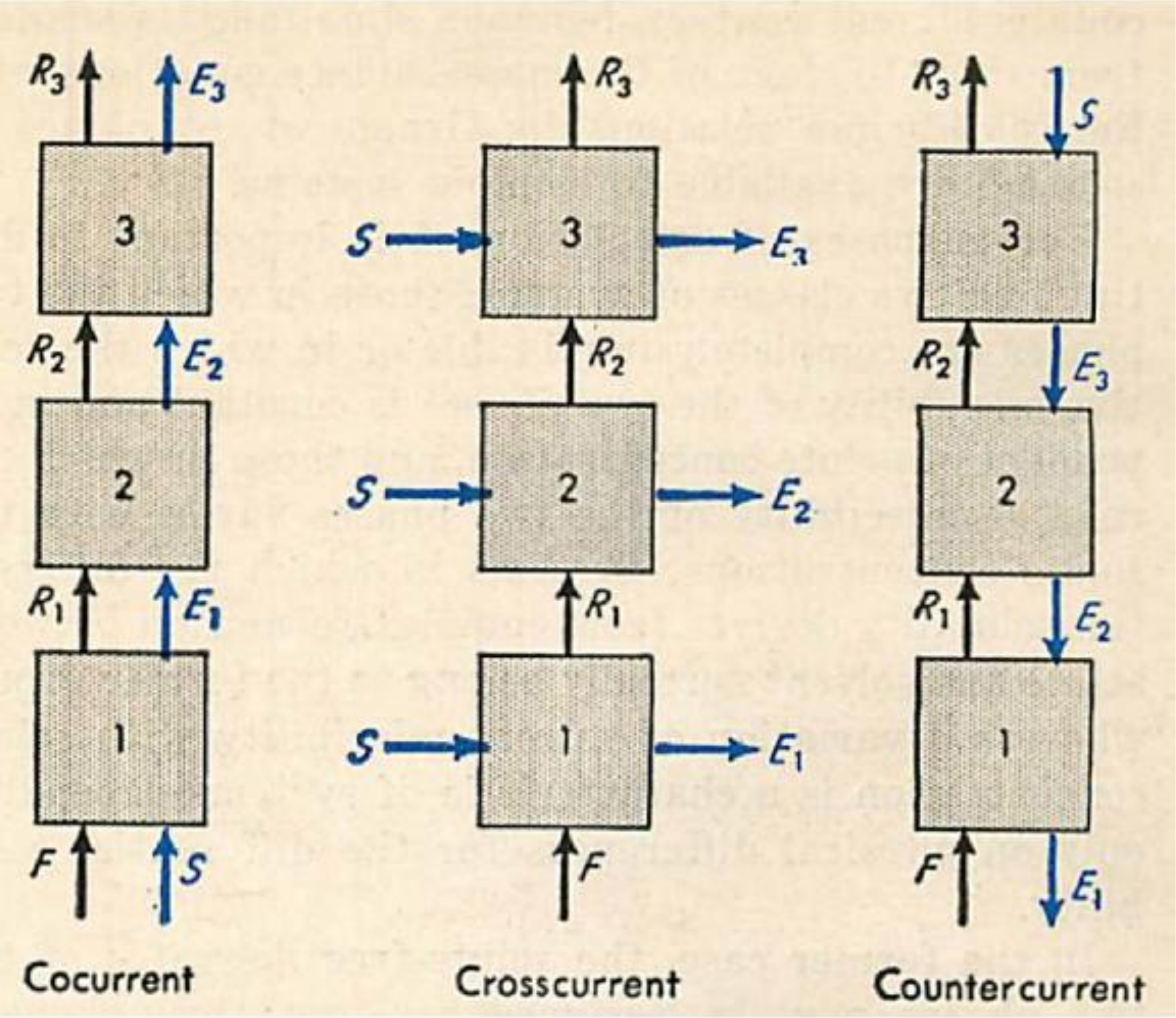
- The choice of solvent for a LLE process can often have a more **significant impact on the process economics** than any other design decision that has to be made.
- **The ideal solvent has**:

1. A high selectivity for the solute relative to the carrier.
2. A high capacity for dissolving the solute, so as to minimize the solvent-to-feed ratio.
3. A minimal solubility in the carrier.
4. A volatility sufficiently different from the solute:
    - recovery of the solvent can be achieved by distillation.
    - the vapor pressure should not be so high that a high extractor pressure is needed.
    - the vapor pressure should not be so low that a high temperature is needed if the solvent is recovered by distillation.

Selecting a solvent for LLE Processes

5. Stability to maximize the solvent life and minimize the solvent make-up requirement.
6. Inertness to permit use of common materials of construction.
7. A low viscosity to promote phase separation, minimize pressure drop, and provide a high solute mass transfer rate.
8. Nontoxic and nonflammable characteristics.
9. Availability at a relatively low cost.
10. Reasonable surface tension: easy to disperse and coalesce.
11. A large difference in density relative to the carrier to achieve a high capacity in the extractor.
12. Compatibility with the solute and carrier to avoid contamination.
13. No tendency to form a stable rag or scum (foam) layer at the phase interface
14. Desirable wetting characteristics with respect to extractor internals

Modes LLE Operations



Some LLE Equipments

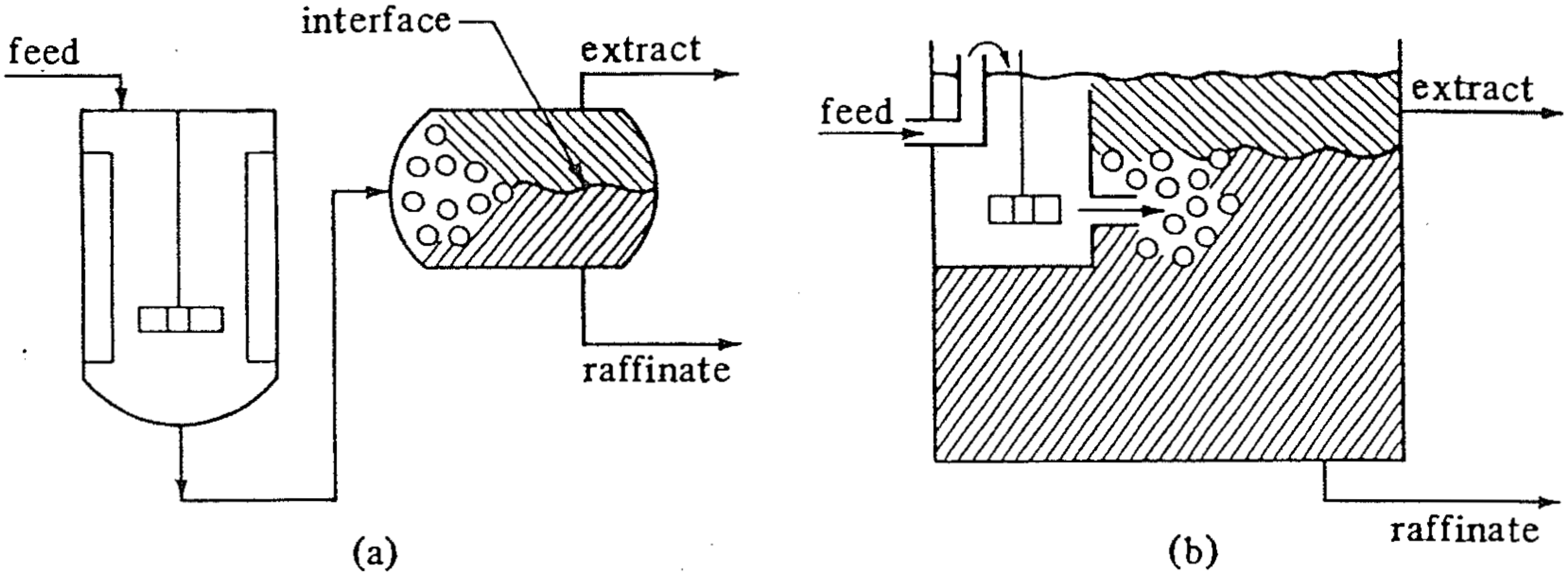


FIGURE 12.6-1. *Typical mixer–settlers for extraction: (a) separate mixer–settler, (b) combined mixer–settler.*

Some LLE Equipments

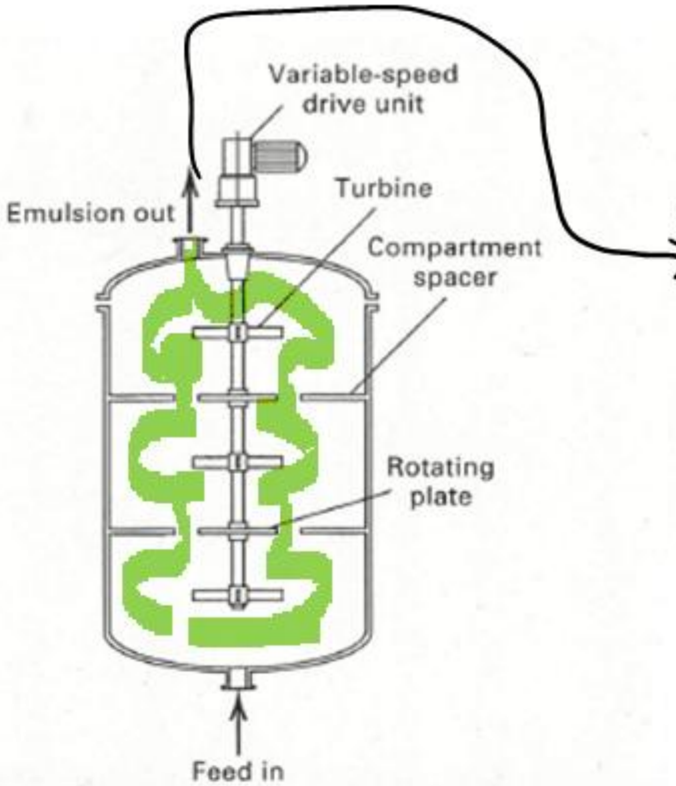


Figure 8.2 Compartmented mixing vessel with variable-speed turbine agitators.

[Adapted from R.E. Treybal, *Mass Transfer,* 3rd ed., McGraw-Hill, New York (1980).]

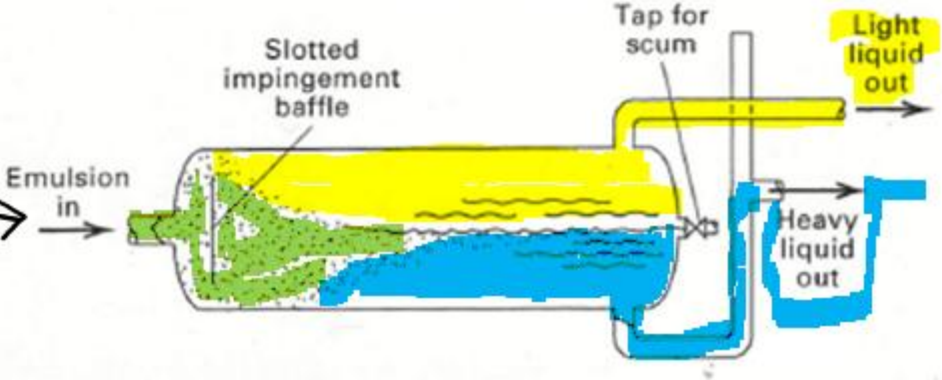


Figure 8.4 Horizontal gravity-settling vessel.

[Adapted from R.E. Treybal, *Liquid Extraction,* 2nd ed., McGraw-Hill, New York (1963) with permission.]

Seader & Henley (2006)

Some LLE Equipments

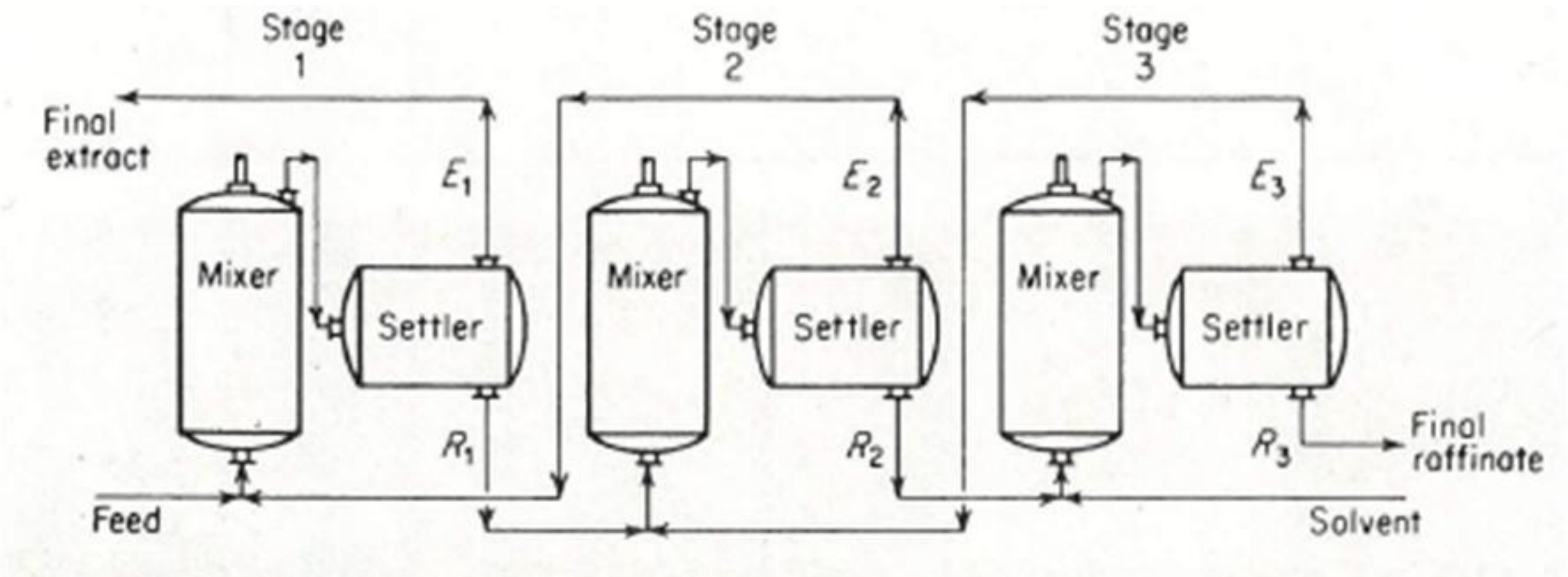
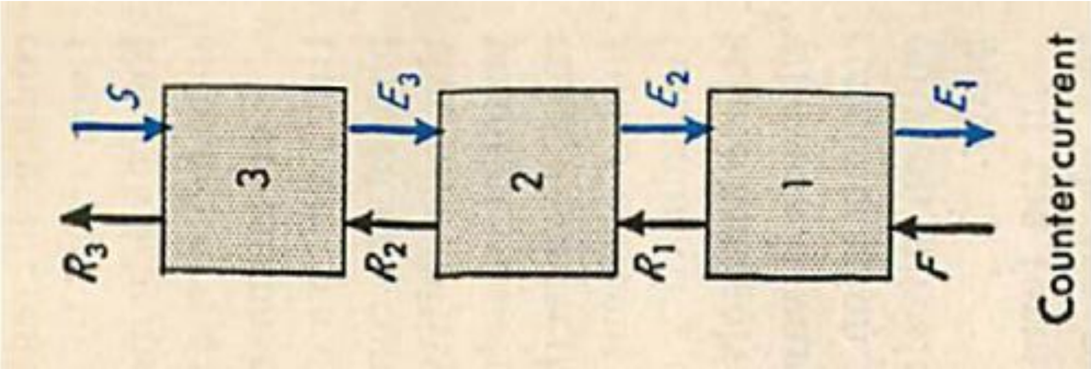


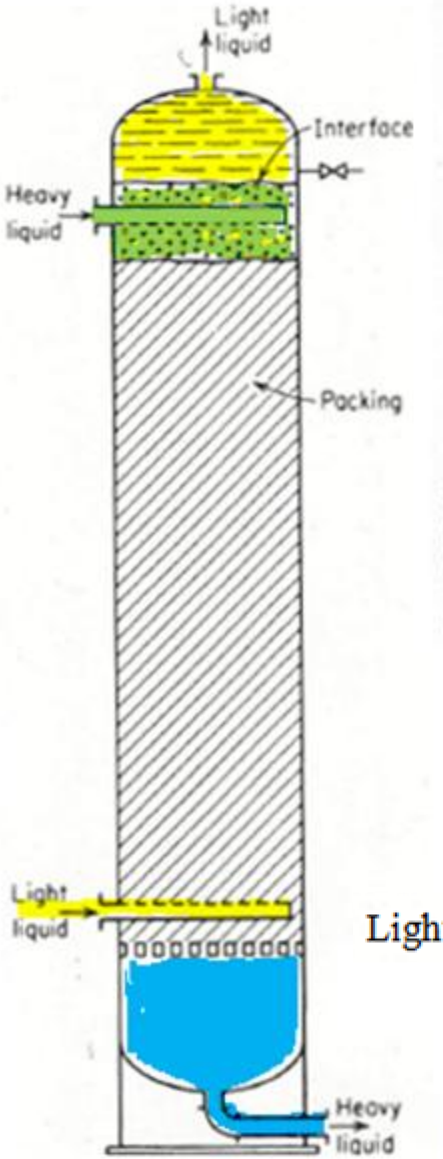
Figure 10.41 Flowsheet of three-stage countercurrent mixer-settler extraction cascade.

Treybal (1980)



Some LLE Equipments

Packed-bed Column



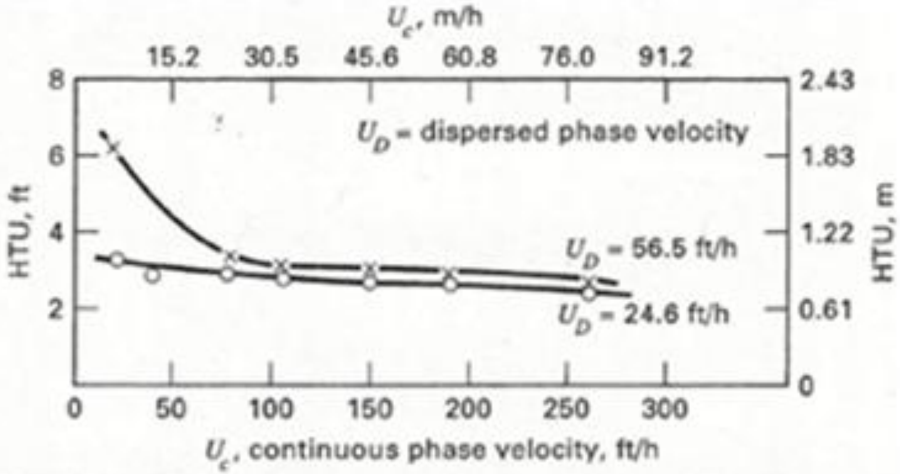


Figure 8.6 Efficiency of 1-in. Intalox saddles in a column 60 in. high with MEK–water–kerosene.

[From R.R. Neumatis, J.S. Eckert, E.H. Foote, and L.R. Rollinson, *Chem. Eng. Progr.*, 67(1), 60 (1971) with permission.]

Light liquid - dispersed phase

Trevbal (1980)

Some LLE Equipments

Spray Columns:

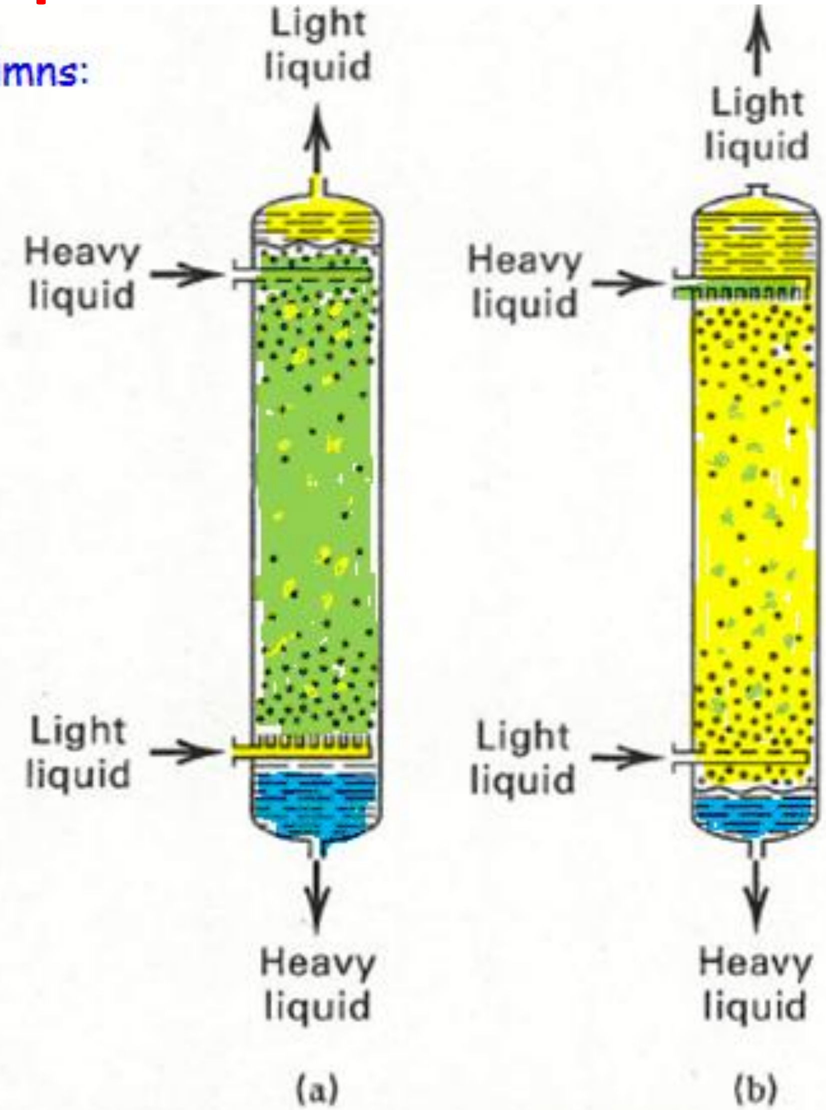


Figure 8.5 Spray columns: (a) light liquid dispersed, heavy liquid continuous; (b) heavy liquid dispersed, light liquid continuous.

Seader & Henley (2006)

Some LLE Equipments

Sieve-tray Extraction Column:
light phase dispersed

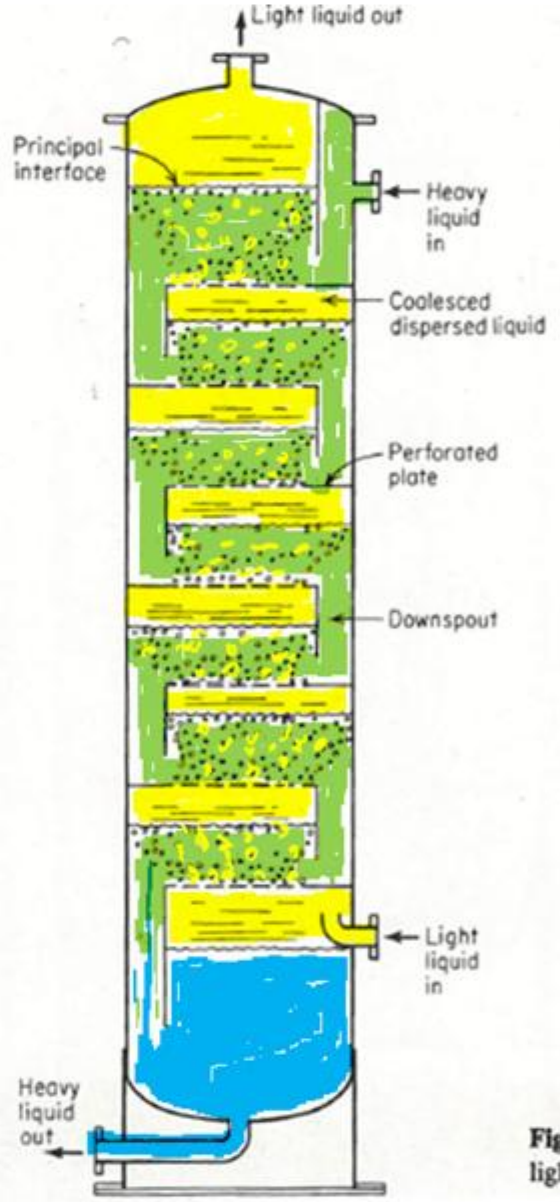
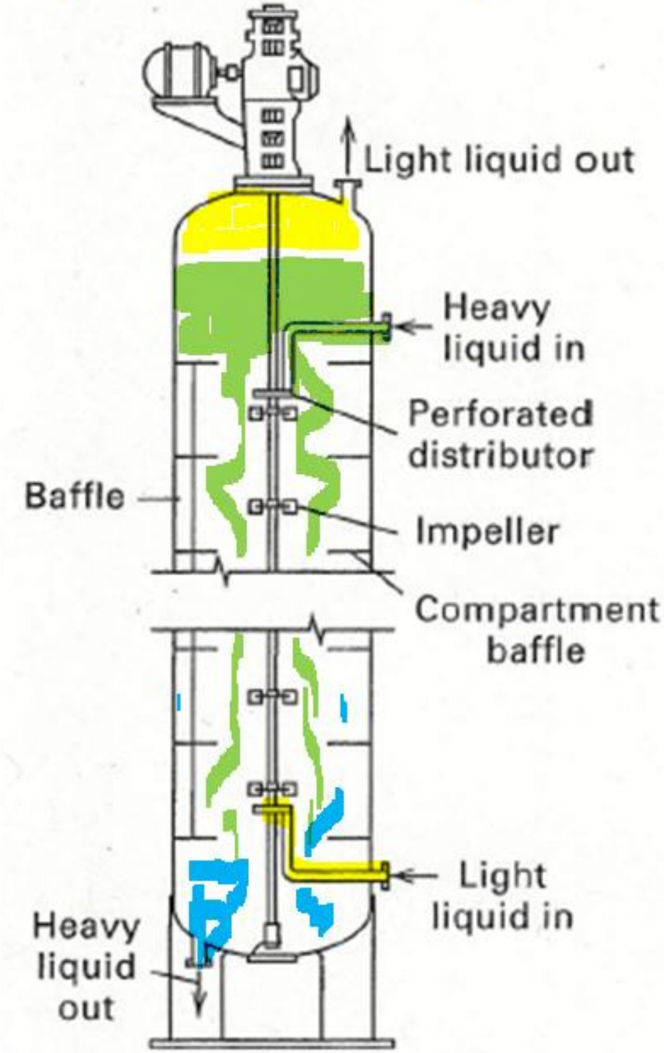


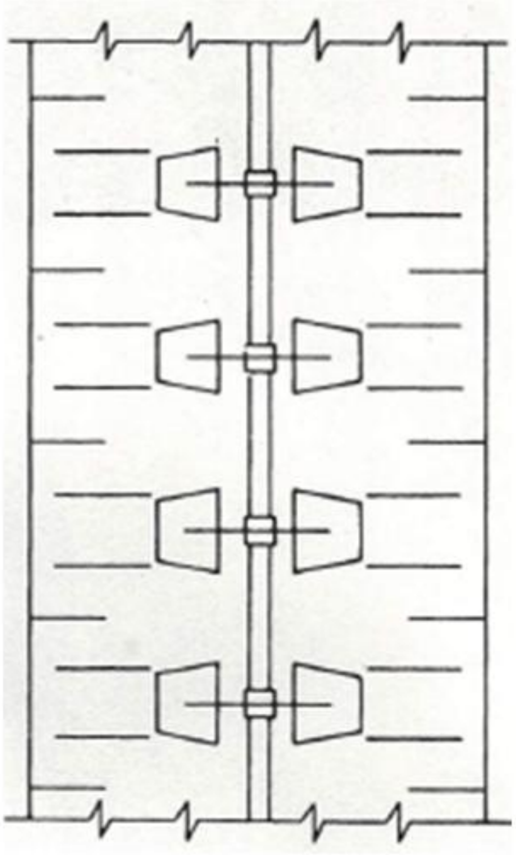
Figure 10.44 Sieve-tray extraction tower, arranged for light liquid dispersed. Treybal (1980)

Some LLE Equipments

Oldshue-Rushton (Mixco Lightnin CMContactor) column



Scheibel column



Seader & Henley (2006)

Some LLE Equipments

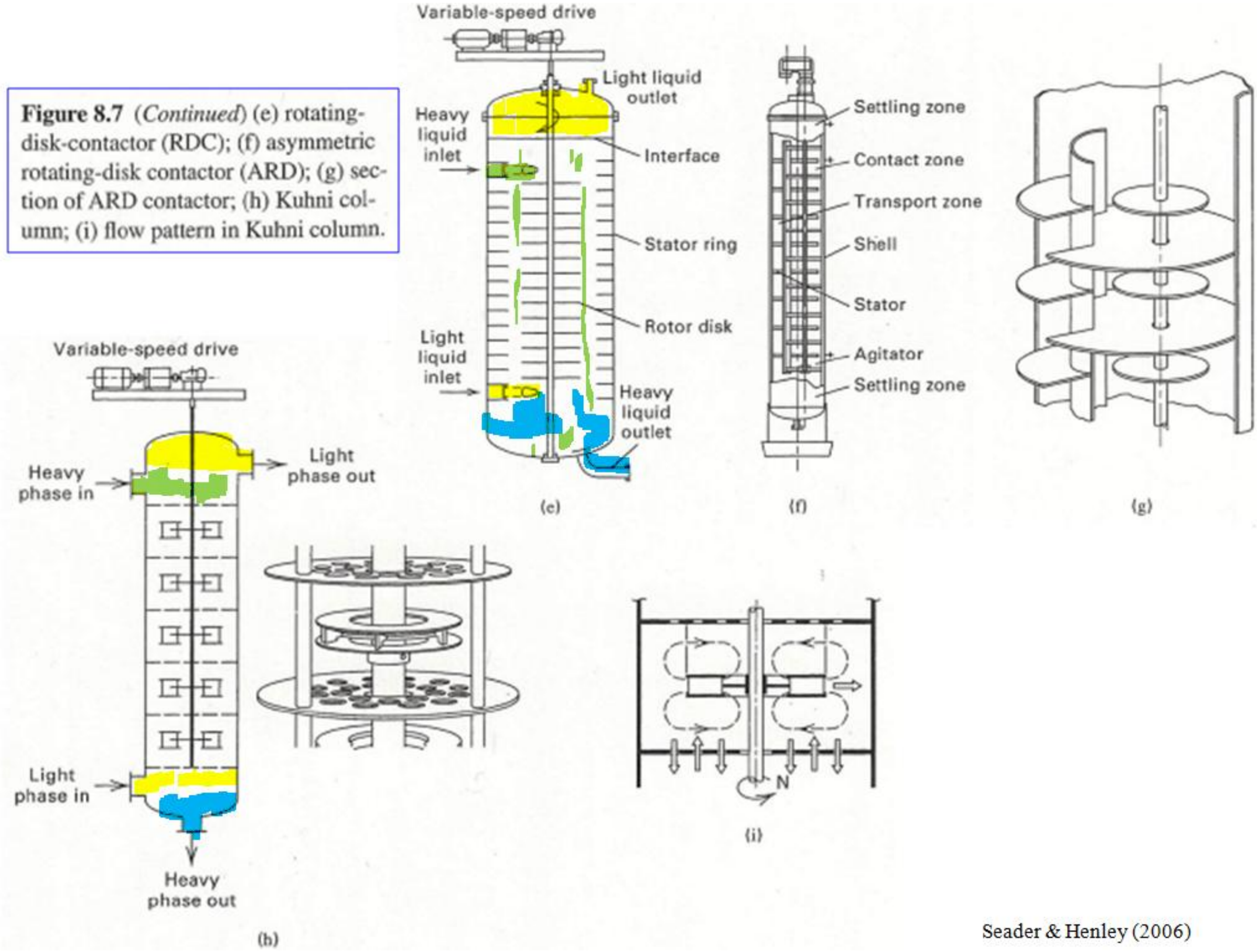


Figure 8.7 (*Continued*) (e) rotating-disk-contactor (RDC); (f) asymmetric rotating-disk contactor (ARD); (g) section of ARD contactor; (h) Kuhni column; (i) flow pattern in Kuhni column.

Seader & Henley (2006)

Some LLE Equipments

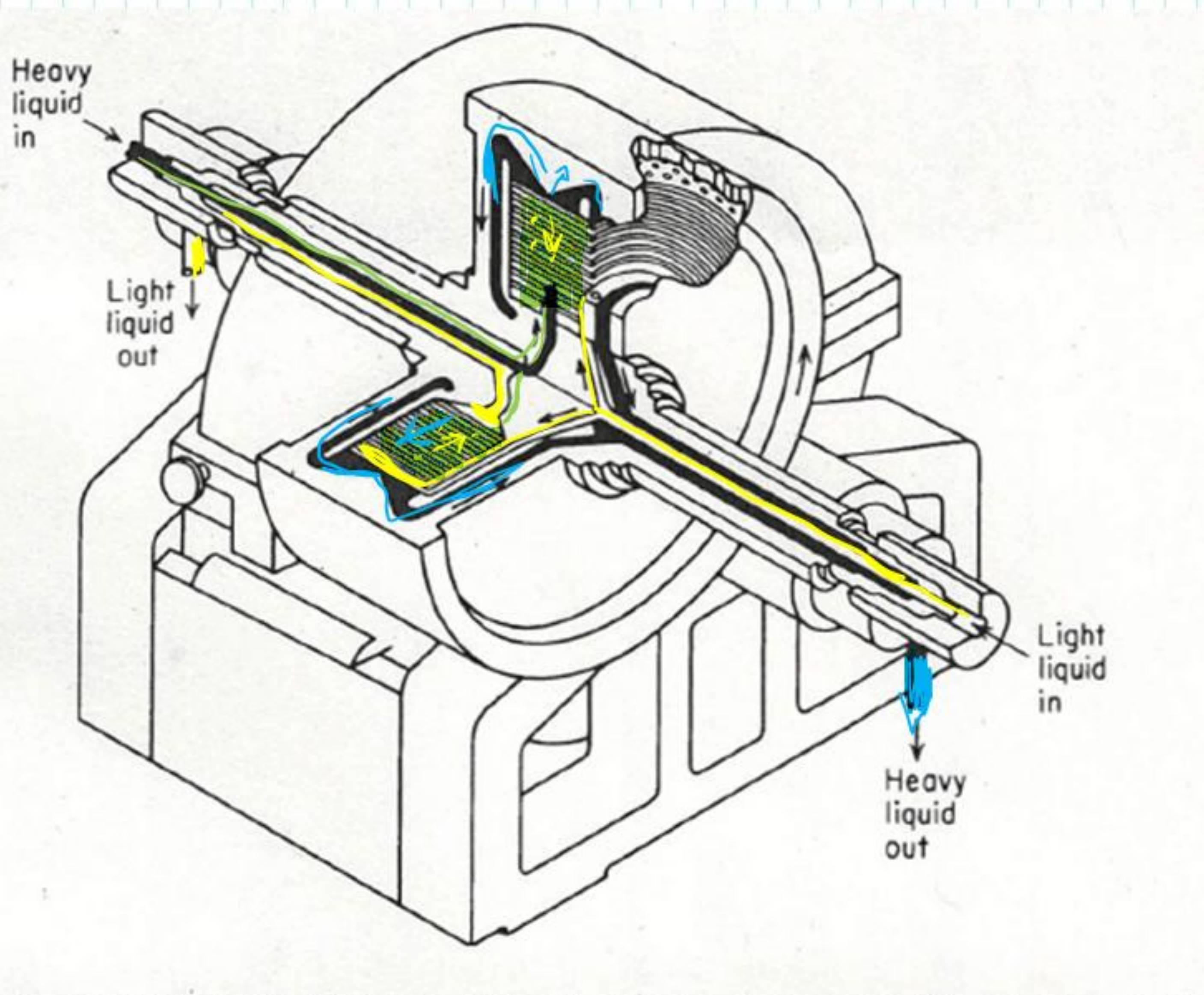


Figure 10.56 Podbielniak centrifugal extractor (schematic). *(Podbielniak, Inc.)*

Some LLE Equipments

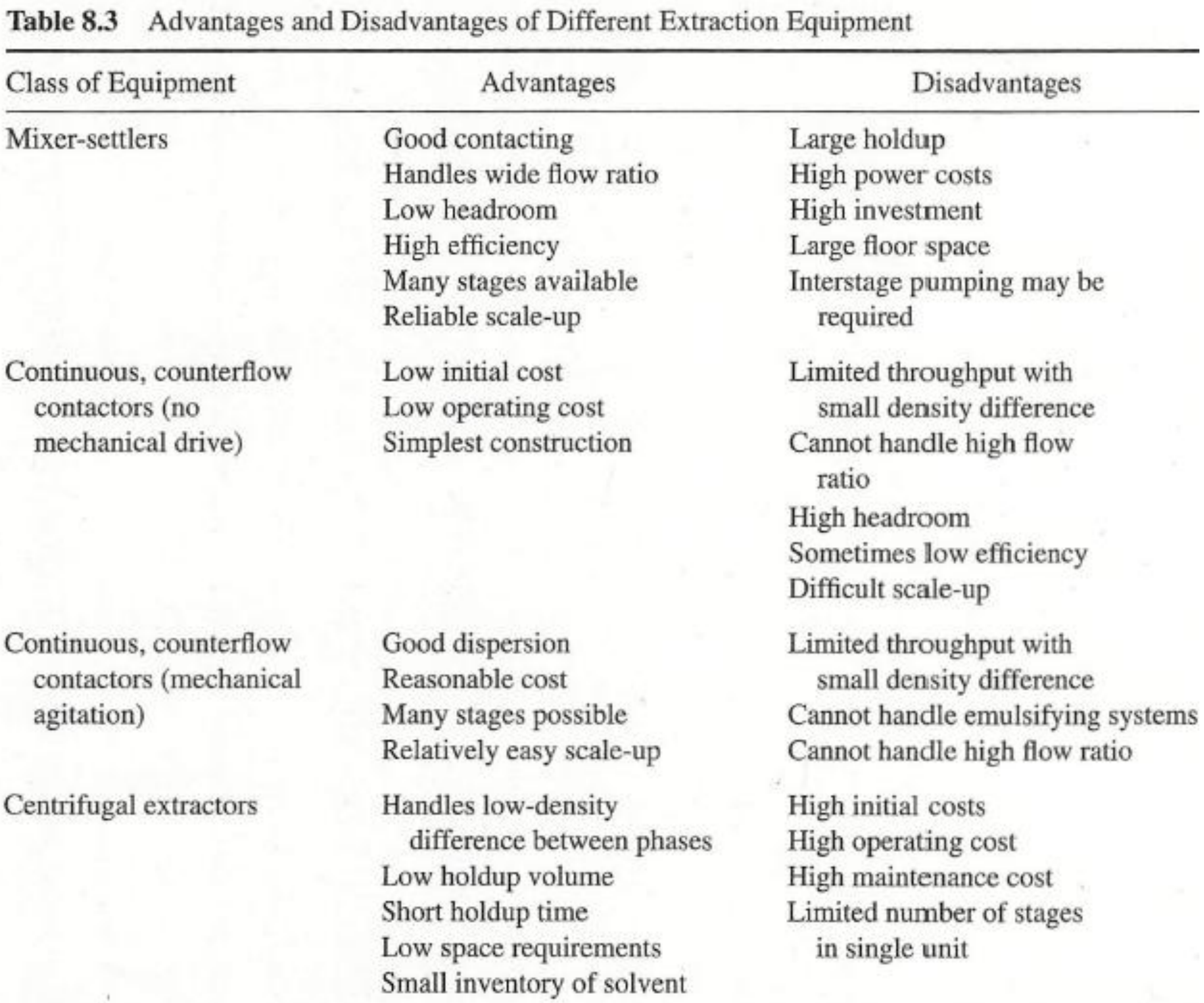


Table 8.3 Advantages and Disadvantages of Different Extraction Equipment

Class of Equipment	Advantages	Disadvantages
Mixer-settlers	Good contacting Handles wide flow ratio Low headroom High efficiency Many stages available Reliable scale-up	Large holdup High power costs High investment Large floor space Interstage pumping may be required
Continuous, counterflow contactors (no mechanical drive)	Low initial cost Low operating cost Simplest construction	Limited throughput with small density difference Cannot handle high flow ratio High headroom Sometimes low efficiency Difficult scale-up
Continuous, counterflow contactors (mechanical agitation)	Good dispersion Reasonable cost Many stages possible Relatively easy scale-up	Limited throughput with small density difference Cannot handle emulsifying systems Cannot handle high flow ratio
Centrifugal extractors	Handles low-density difference between phases Low holdup volume Short holdup time Low space requirements Small inventory of solvent	High initial costs High operating cost High maintenance cost Limited number of stages in single unit

Some LLE Equipments

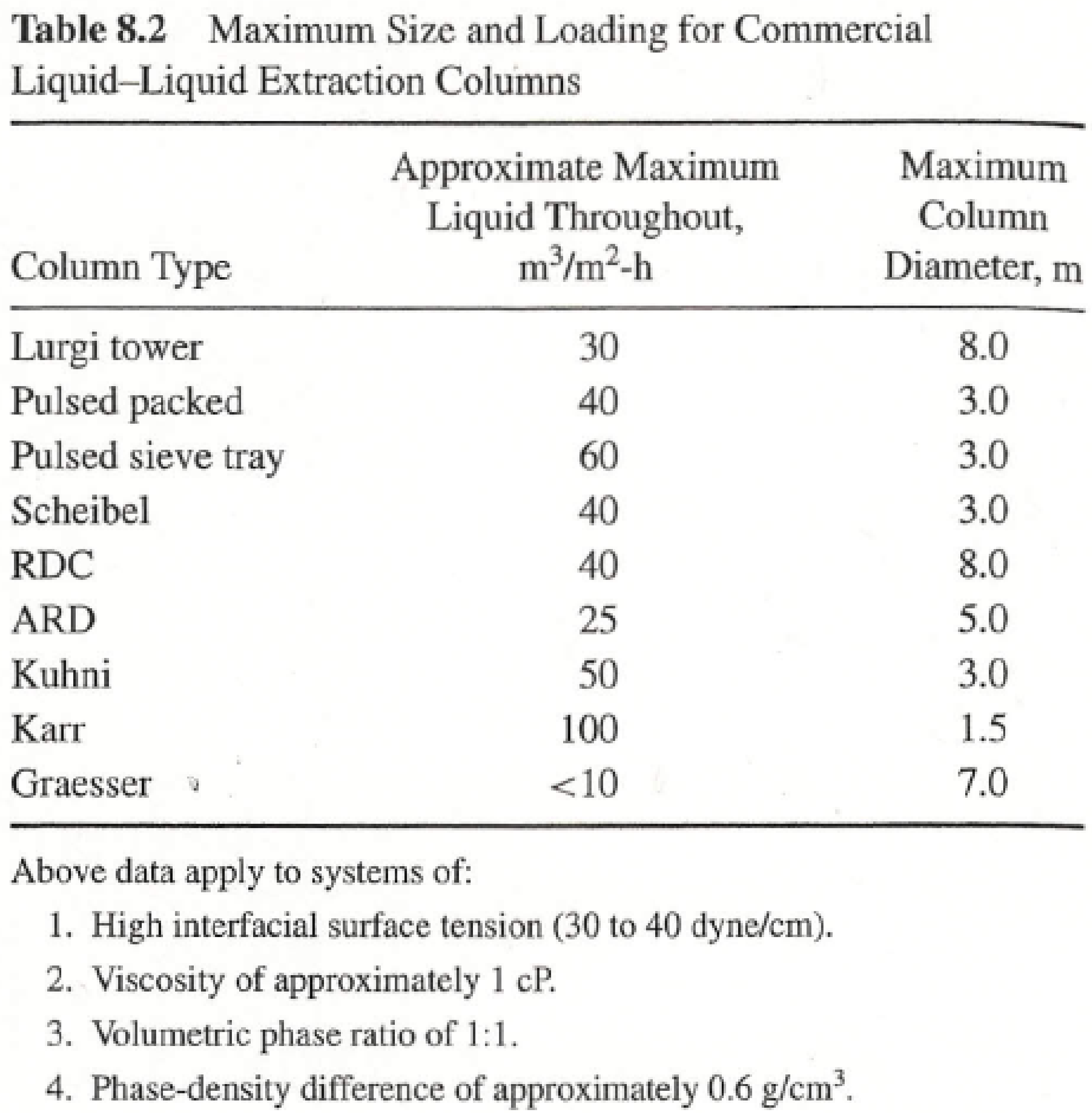


Table 8.2 Maximum Size and Loading for Commercial Liquid–Liquid Extraction Columns

Column Type	Approximate Maximum Liquid Throughout, m^3/m^2-h	Maximum Column Diameter, m
Lurgi tower	30	8.0
Pulsed packed	40	3.0
Pulsed sieve tray	60	3.0
Scheibel	40	3.0
RDC	40	8.0
ARD	25	5.0
Kuhni	50	3.0
Karr	100	1.5
Graesser	<10	7.0

Above data apply to systems of:

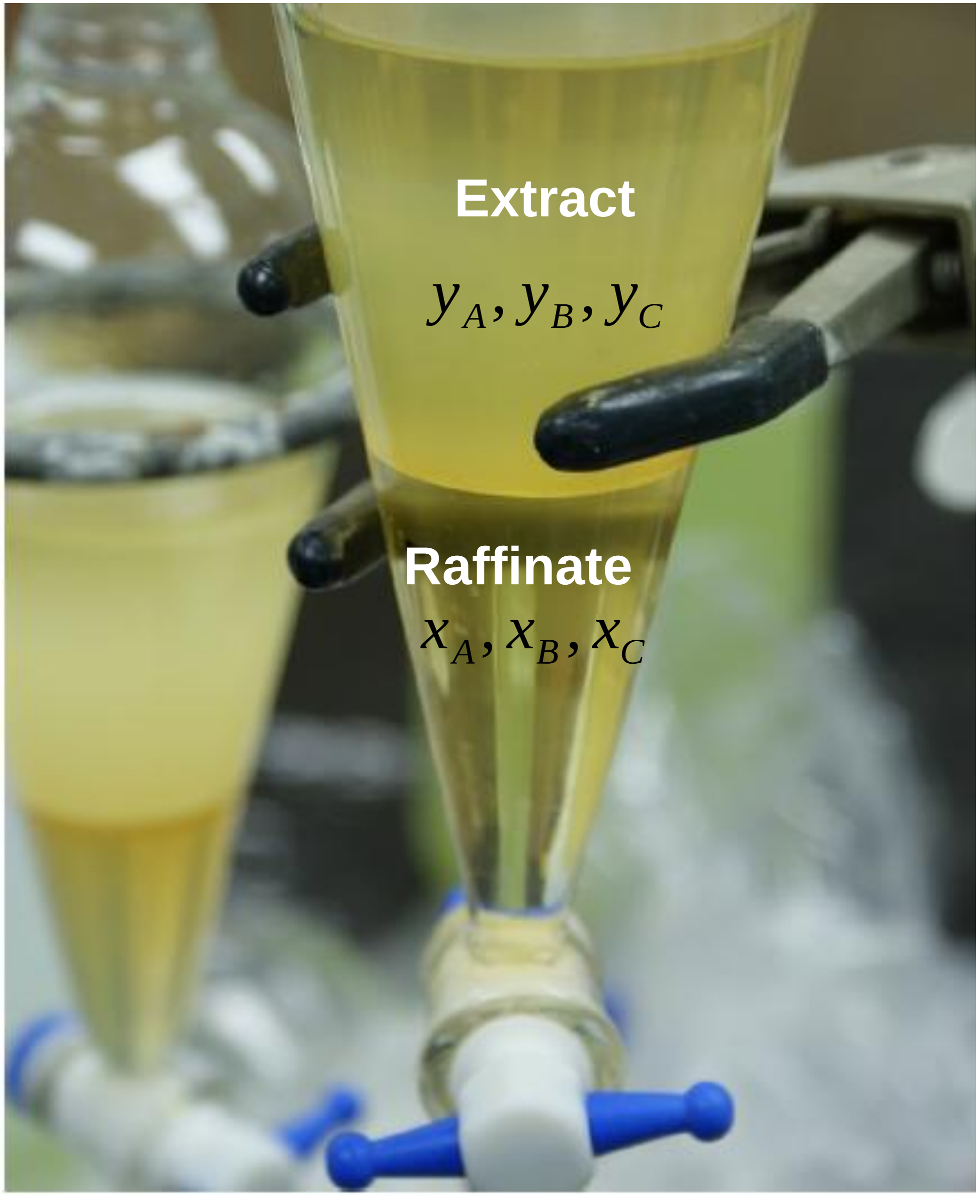
1. High interfacial surface tension (30 to 40 dyne/cm).
2. Viscosity of approximately 1 cP.
3. Volumetric phase ratio of 1:1.
4. Phase-density difference of approximately 0.6 g/cm^3.

Liquid-liquid equilibrium in extraction

- **Two phases:**
  - Extract
  - Raffinate phases

- **Usually three components:**
  - Solute (A)
  - Carrier (B)
  - Solvent (C or S)

  (ternary system)



At certain Temperature and pressure

Liquid-liquid equilibrium in extraction

Gibbs phase rule:

$$\boxed{F = 2 - \pi + N}$$

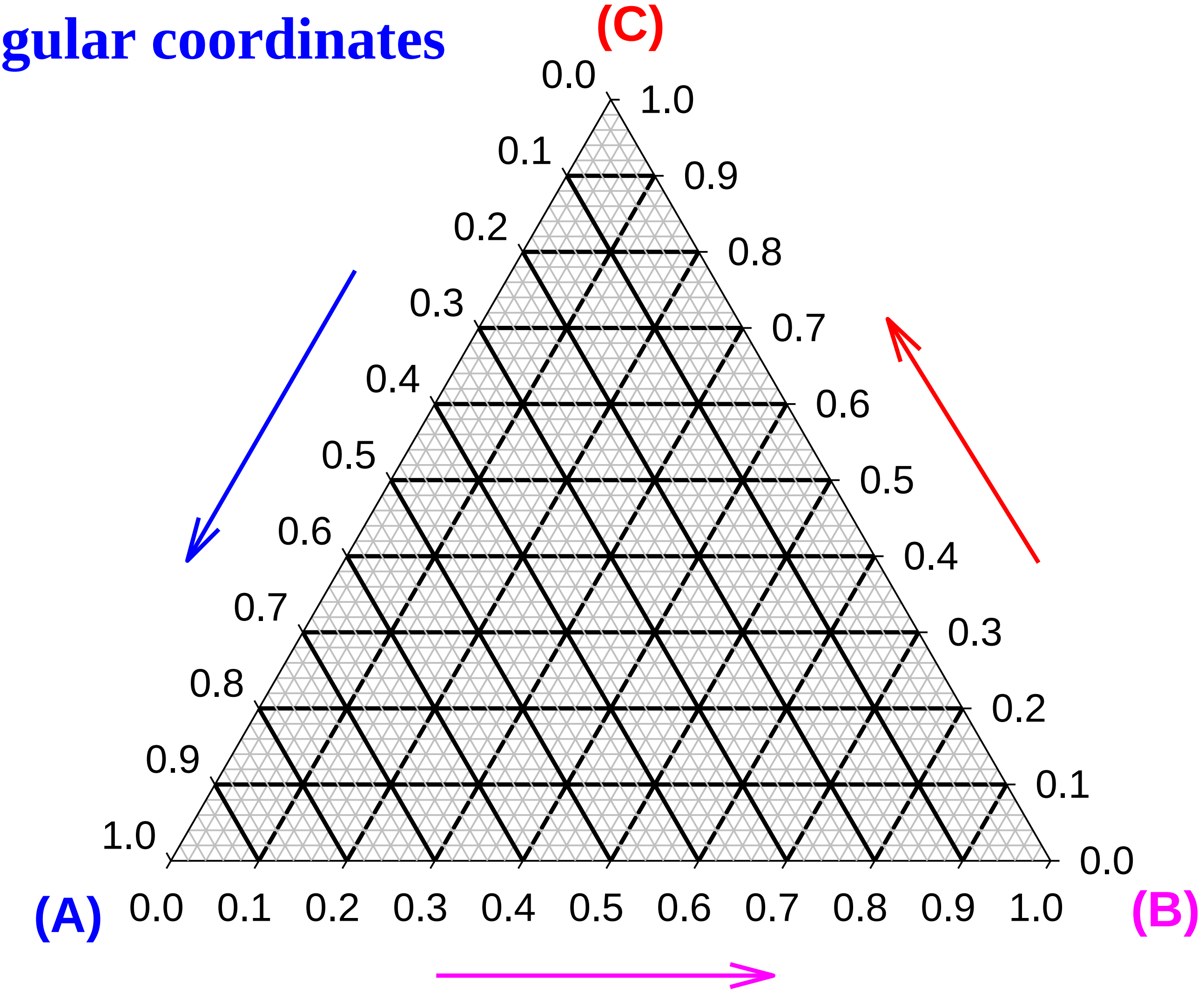
where F is the degrees of freedom for nonreacting systems:
π is the number of phases, N is the number of chemical species.

- For LLE: F=2-2+3=3 → three intensive properties can be specified and the others can be determined.
- **Usual case**: temperature, pressure and one species concentration in either extract or raffinate phase are set. At equilibrium, the other five concentrations can be determined.
- Remember that:

$$\boxed{\begin{array}{ll} x_A + x_B + x_C = 1 & (Raffinate\ phase) \\ y_A + y_B + y_C = 1 & (Extract\ \ phase) \end{array}}$$

Liquid-Liquid Equilibrium Phase Diagram

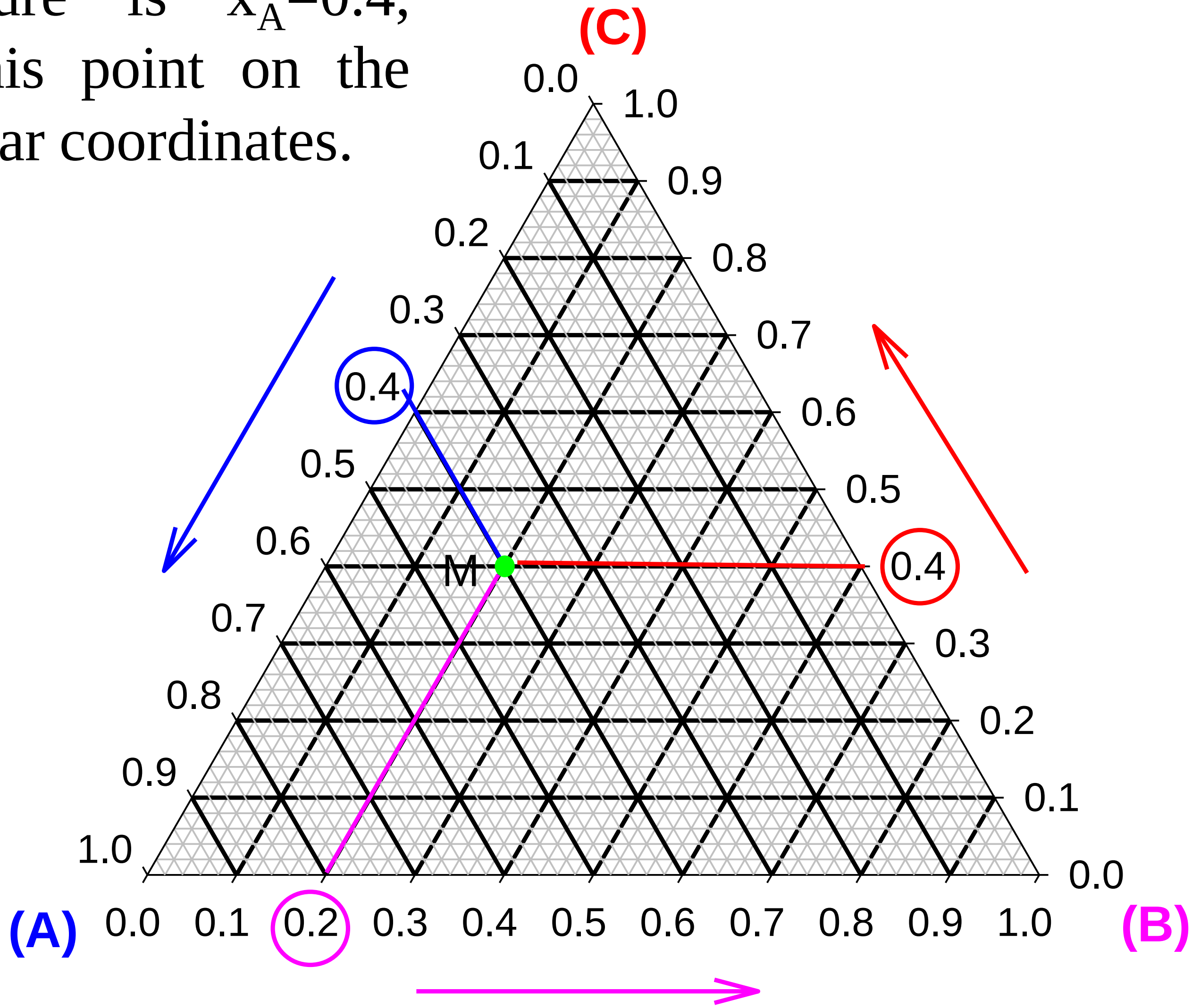
Equilateral triangular coordinates



Liquid-Liquid Equilibrium Phase Diagram

Example. The mass composition of ternary mixture is x_A=0.4, x_B=0.2. Locate this point on the equilateral triangular coordinates.

$x_c = 1 - x_A - x_B = 0.4$

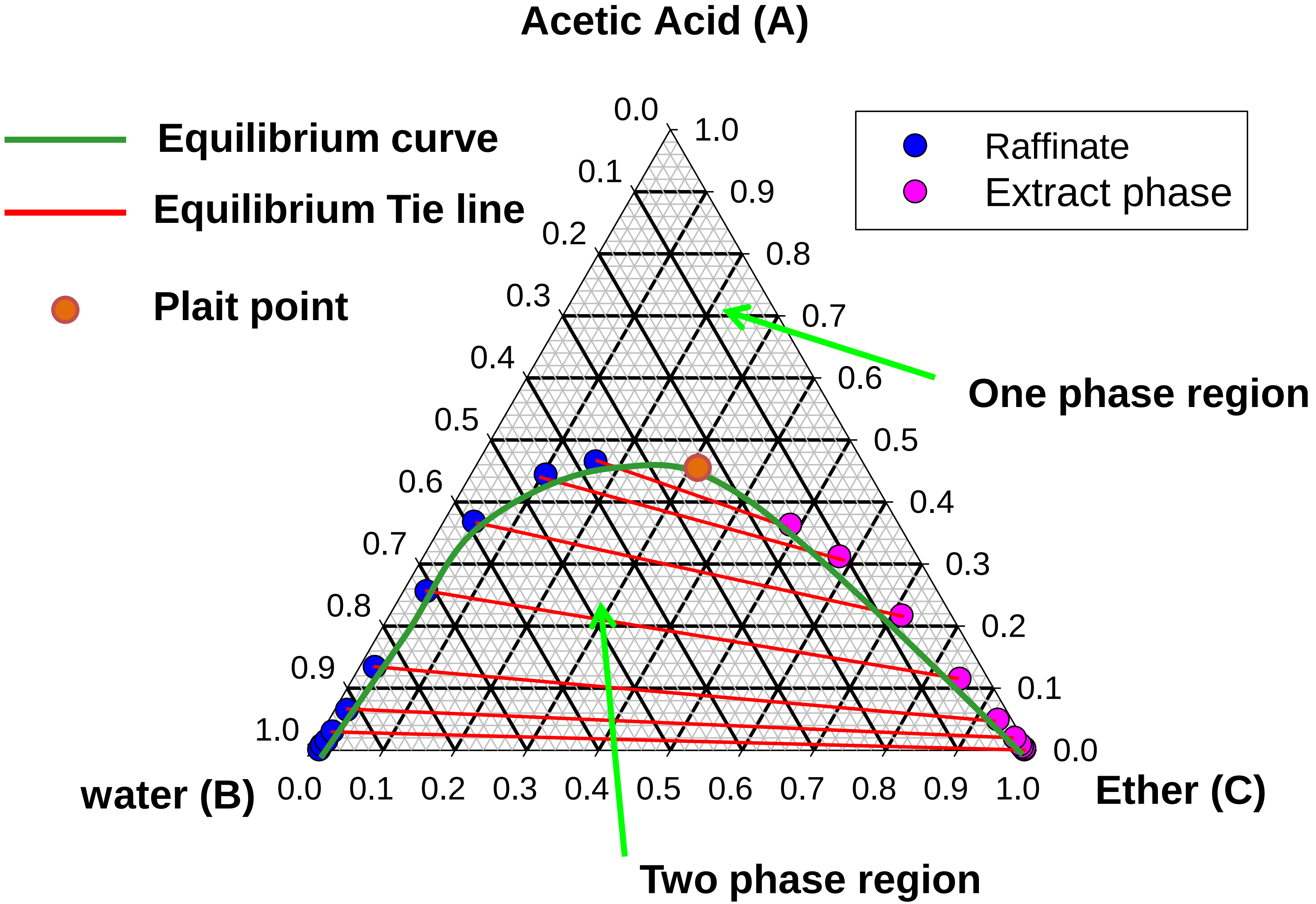


Liquid-Liquid Equilibrium Phase Diagram

Example. Given liquid-liquid equilibrium data for acetic acid (solute)-water (carrier)- Isopropyl ether (solvent) at 20 ^{o}C and 1 atm. Draw the phase diagram on the equilateral triangular coordinates.

	Mass fraction (x) water layer (Raffinate)			Mass fraction (y) Isopropyl ether layer (extract)		
Run	**Acetic acid**	**Water**	**Isopropyl Ether**	**Acetic acid**	**Water**	**Isopropyl Ether**
1	0.0000	0.9880	0.0120	0.0000	0.0060	0.9940
2	0.0070	0.9810	0.0120	0.0020	0.0050	0.9930
3	0.0140	0.9710	0.0150	0.0040	0.0070	0.9890
4	0.0290	0.9550	0.0160	0.0080	0.0080	0.9840
5	0.0640	0.9170	0.0190	0.0190	0.0100	0.9710
6	0.1330	0.8440	0.0230	0.0480	0.0190	0.9330
7	0.2550	0.7110	0.0340	0.1140	0.0390	0.8470
8	0.3670	0.5890	0.0440	0.2160	0.0690	0.7150
9	0.4430	0.4510	0.1060	0.3110	0.1080	0.5810
10	0.4640	0.3710	0.1650	0.3620	0.1510	0.4870

Liquid-Liquid Equilibrium Phase Diagram



Equilibrium curve

Equilibrium Tie line

Plait point

One phase region

Two phase region

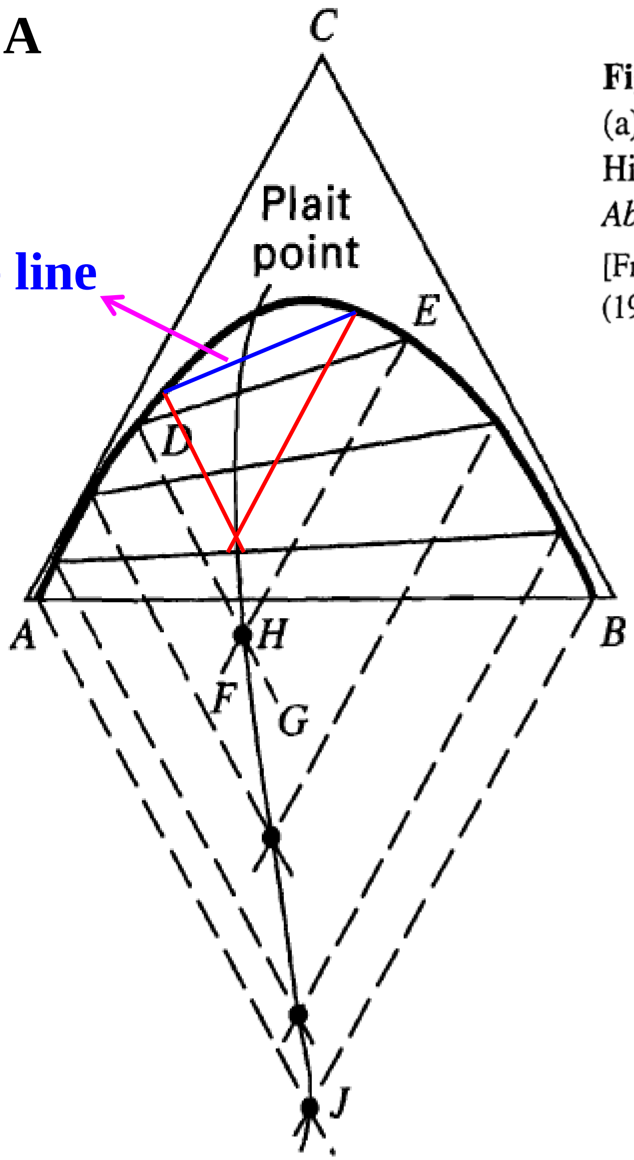
Liquid-Liquid Equilibrium Phase Diagram

- **Tie line:** line connect the raffinate and extract at equilibrium.
- Tie line located within the two phase region

- **Plait point:** point on the equilibrium curve at which the compositions in raffinate and extract phases are identical.
- It is evident in the previous figure that the solvent pair B and C are **partially miscible** (two phase region).

Tie line interpolation

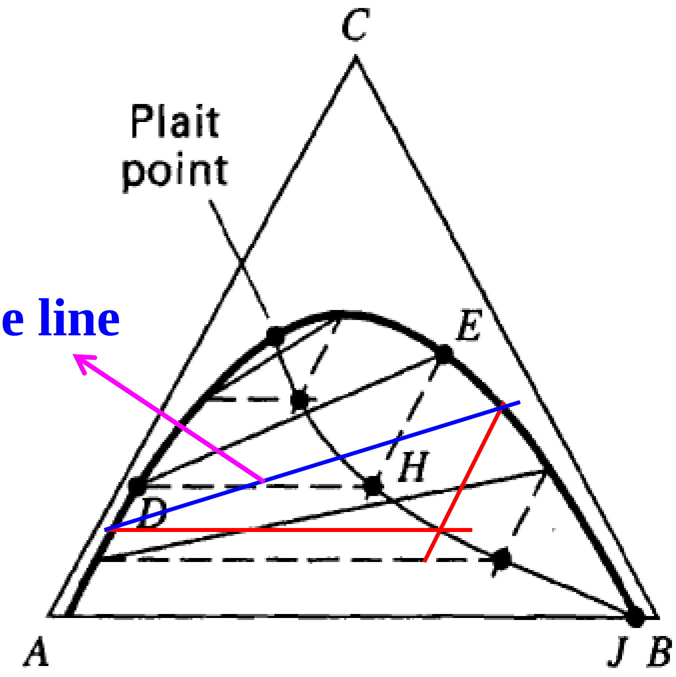
Method A



New tie line

Figure 8.16 Use of conjugate curves to interpolate tie lines: (a) method of *International Critical Tables,* Vol. III, McGraw-Hill, New York, p. 393 (1928); (b) method of T.K. Sherwood, *Absorption and Extraction,* McGraw-Hill, New York, p. 242 (1937). [From R.E. Treybal, *Liquid Extraction,* 2nd ed., McGraw-Hill, New York (1963) with permission.]

Method B



New tie line

Phase diagram on rectangular coordinates

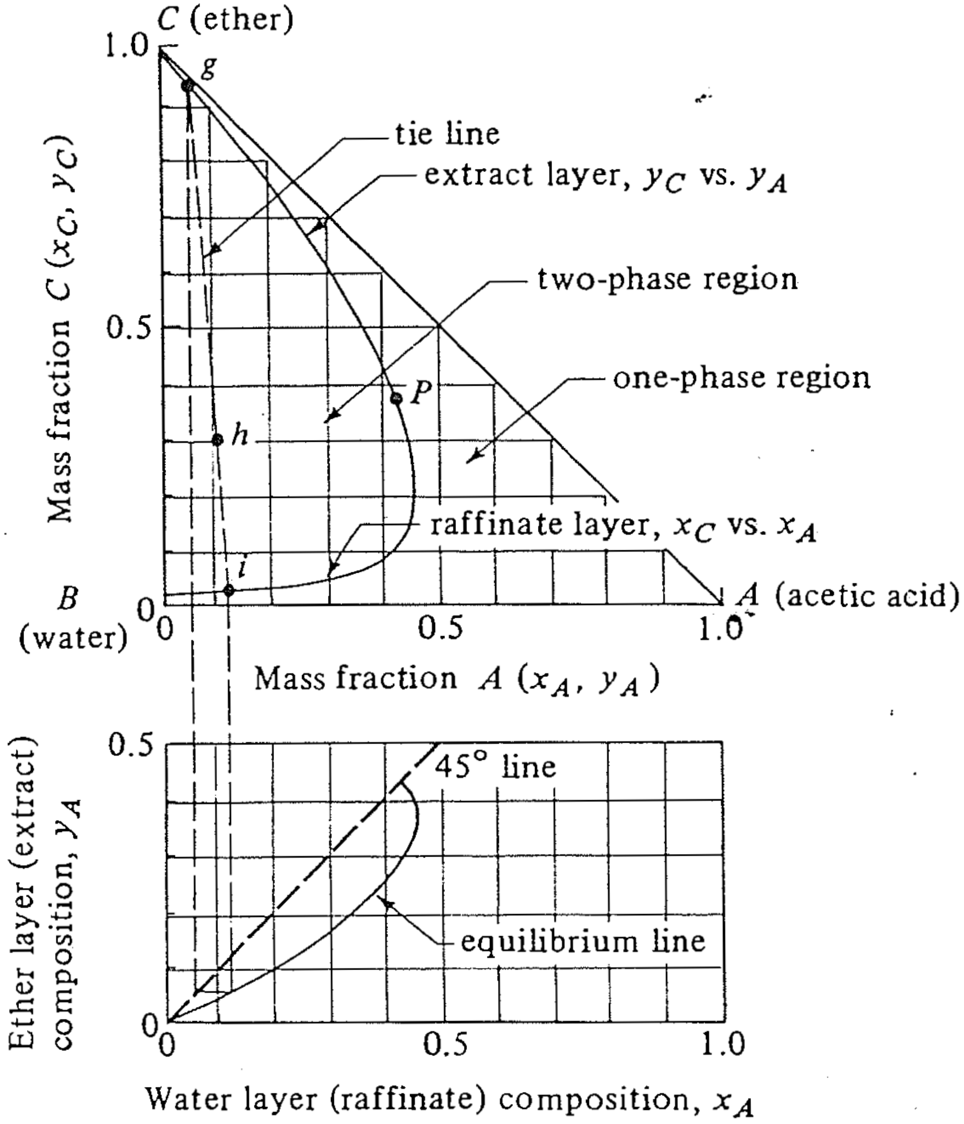


FIGURE 12.5-3. *Acetic acid (A)–water (B)–isopropyl ether (C) liquid–liquid phase diagram at 293 K (20°C).*

Common classes of ternary liquid-liquid systems

- **One immiscible or partially miscible pair:**

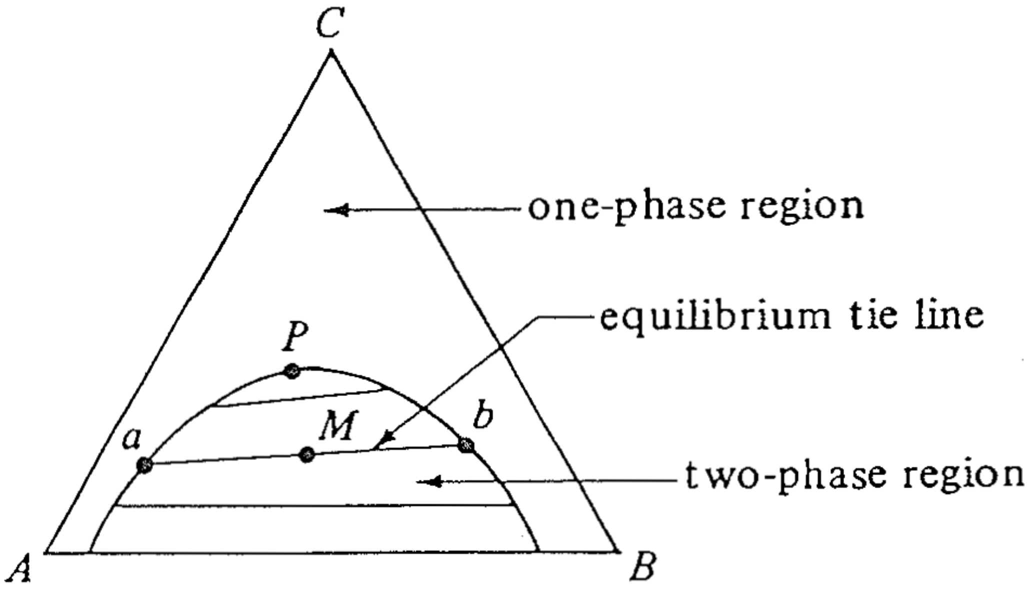


FIGURE 12.5-2. *Liquid–liquid phase diagram where components A and B are partially miscible.*

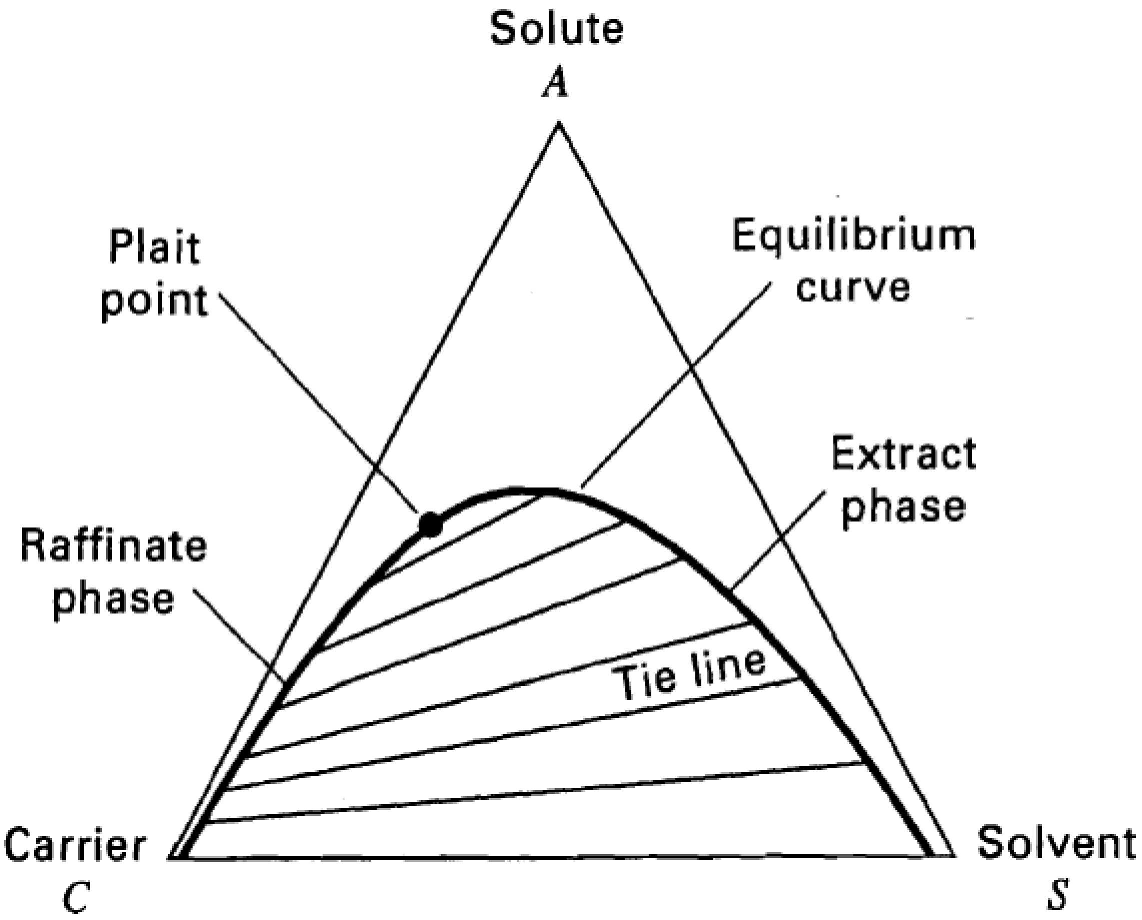
- **In the above phase LL phase diagram:**

- The pair of components of A and B are partially miscible.
- The pair of components A and C are completely miscible.
- The pair of components B and C are completely miscible.

Example: Water(A)-Chloroform(B)-Acetone(C)

Common classes of ternary liquid-liquid systems

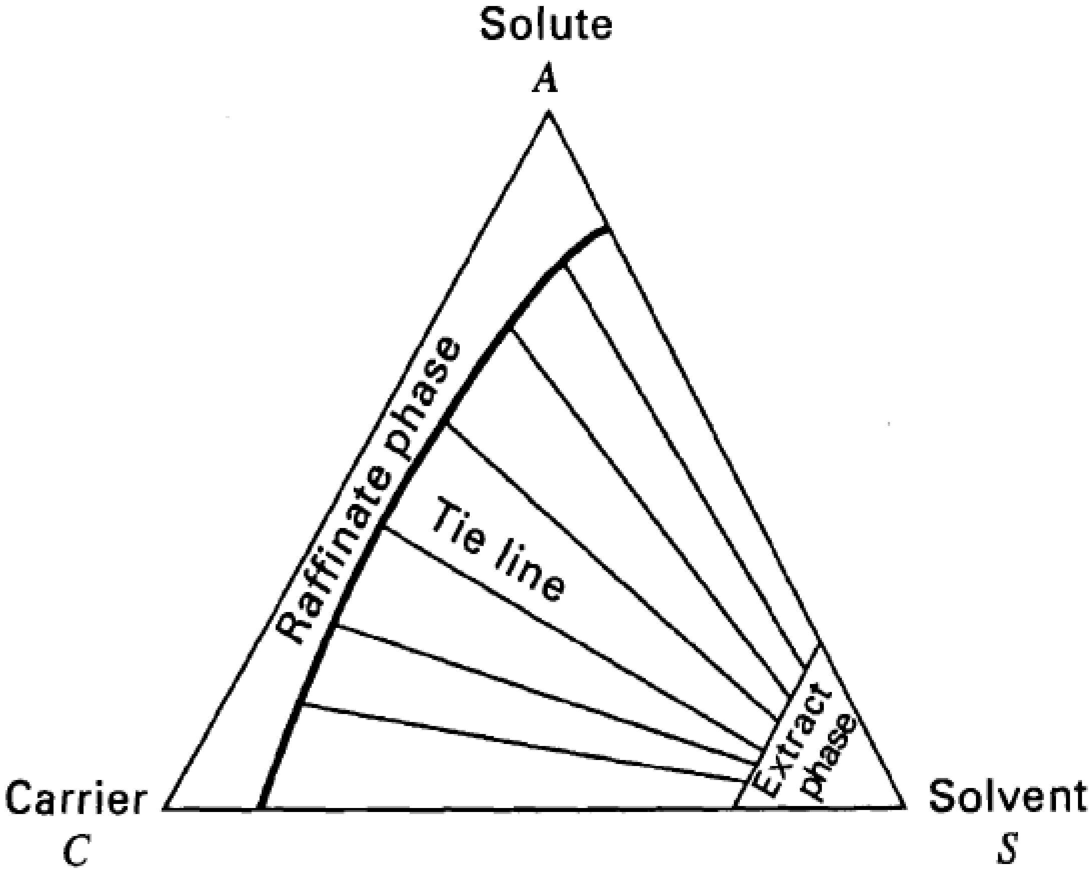
- **One immiscible or partially miscible pair:**



C-S pair is partially miscible

Common classes of ternary liquid-liquid systems

Two immiscible or partially miscible pairs

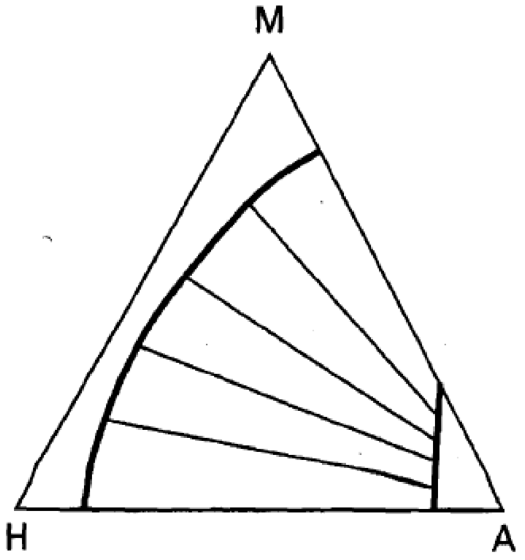


C-S and A-S pairs are partially miscible pair

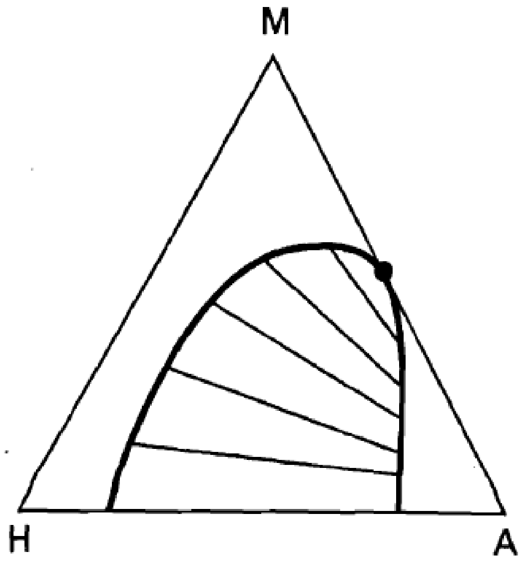
C-A is completely miscible

Effect of temperature on liquid-liquid phase diagram

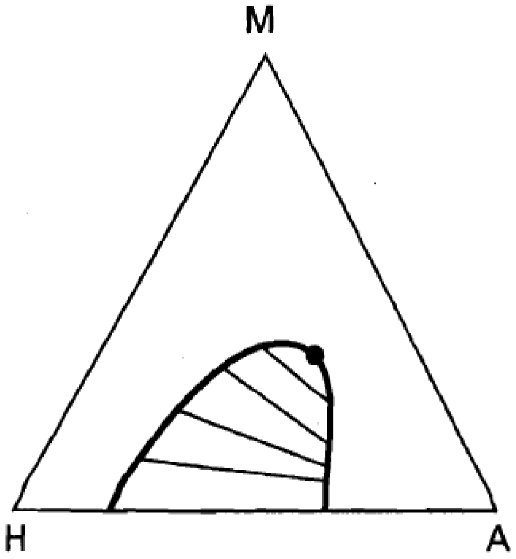
As shown in the figure below, phase diagram varies with temperature variation:



$T = 25°C$



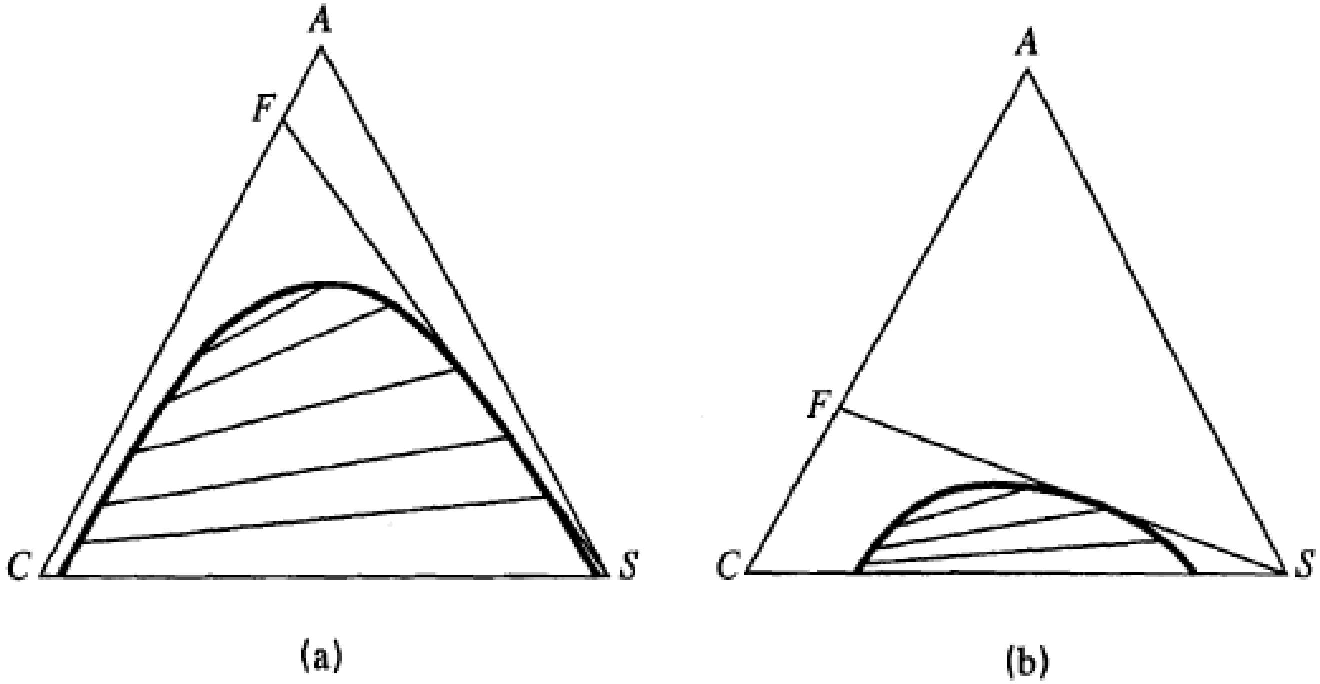
$T = 34.5°C$



$T = 45°C$

Phase diagram and the range of feed composition that can be extracted

Example:



(a) (b)

Figure 8.11 Effect of solubility on range of feed composition that can be extracted.

Phase diagram in (a) has a wider range of feed composition than that of (b)

Phase diagram on rectangular coordinates

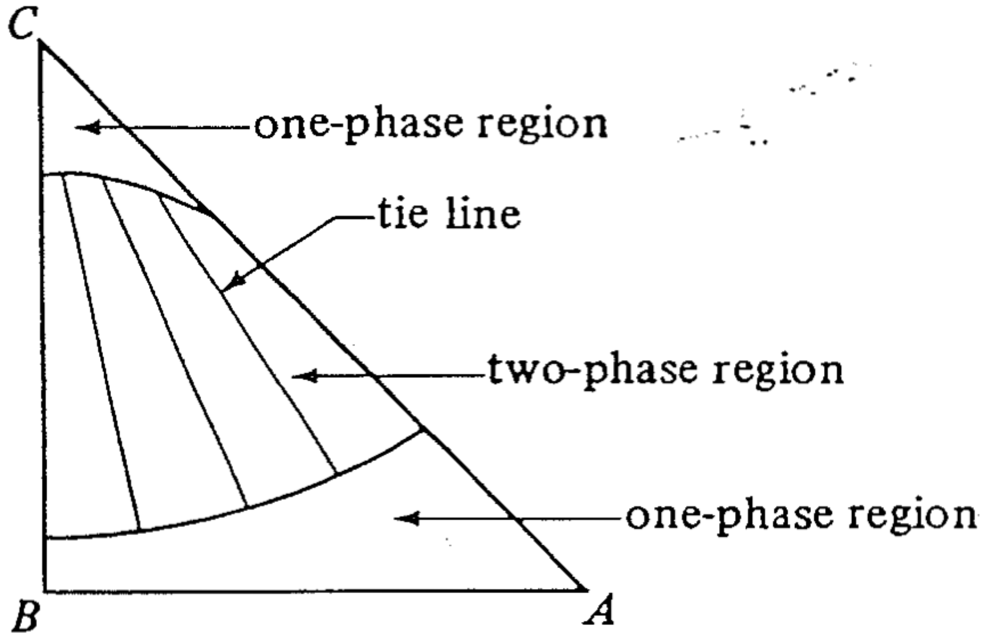


FIGURE 12.5-4. *Phase diagram where the solvent pairs B–C and A–C are partially miscible.*

Lever-arm Rule

- Total mass balance:

$$R+E=M$$

- Component A MB:

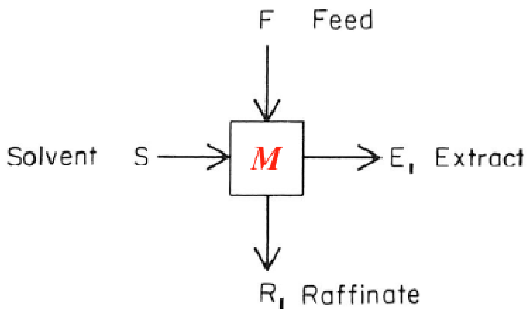
$$Rx_A+Ey_A=Mx_{AM}$$

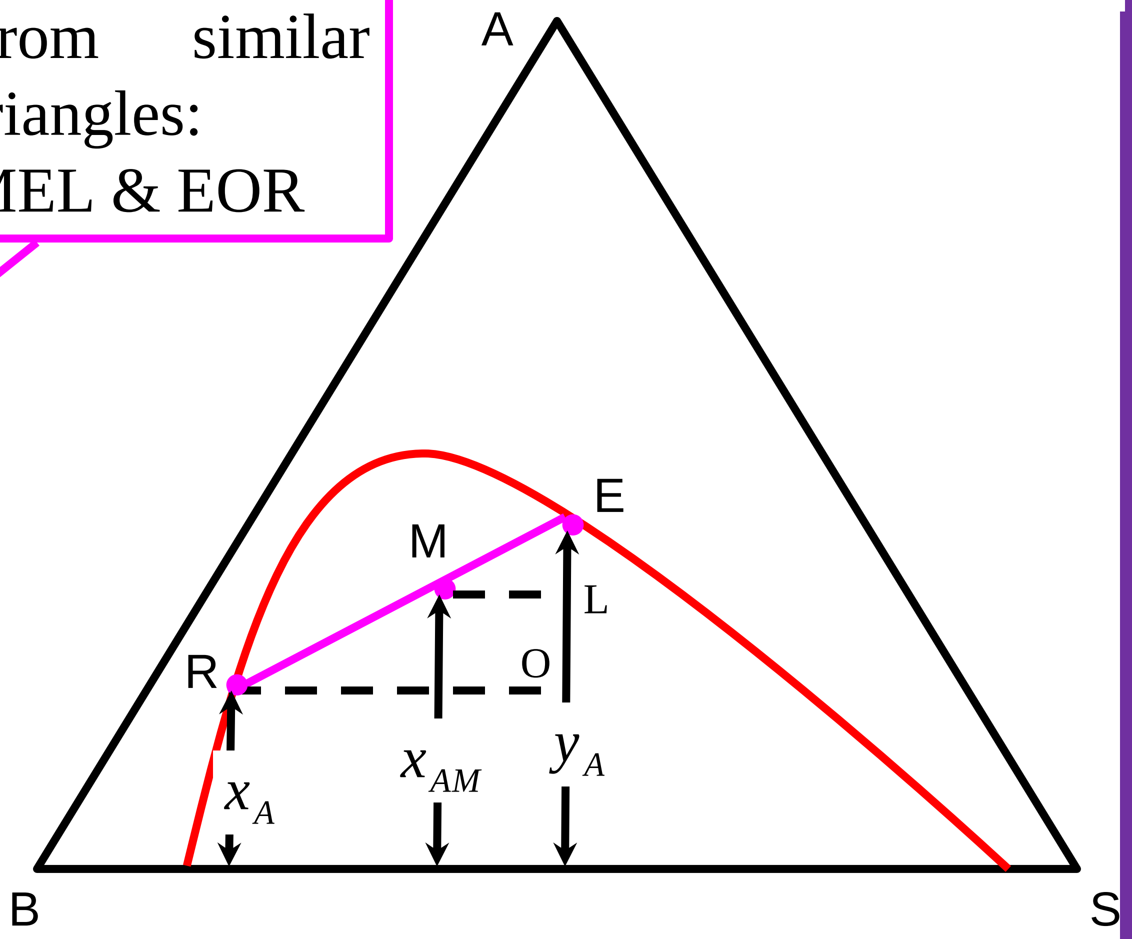
From similar triangles: MEL & EOR

$$\frac{R}{E}=\frac{y_A-x_{AM}}{x_{AM}-x_A}=\frac{\overline{EM}}{\overline{RM}}$$

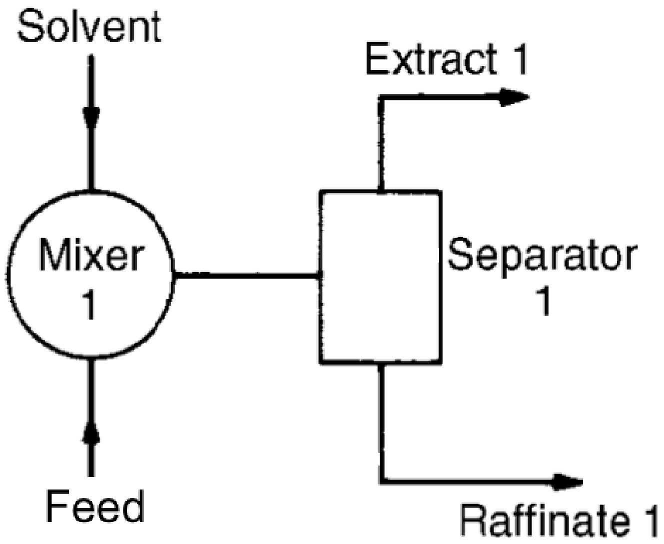
$$\frac{R}{M}=\frac{y_A-x_{AM}}{x_{AM}-x_A}=\frac{\overline{EM}}{\overline{RE}}$$

$$\frac{E}{M}=\frac{y_A-x_{AM}}{x_{AM}-x_A}=\frac{\overline{RM}}{\overline{RE}}$$





One-stage extraction calculations



Example. In mixer-settler extraction unit, 250 kg of feed which contains 24 wt% solute(A) 76 wt% carrier(C) is mixed with pure 100 kg solvent (S).

1) Find the overall composition of mixture at equilibrium using:

a) The phase diagram given below

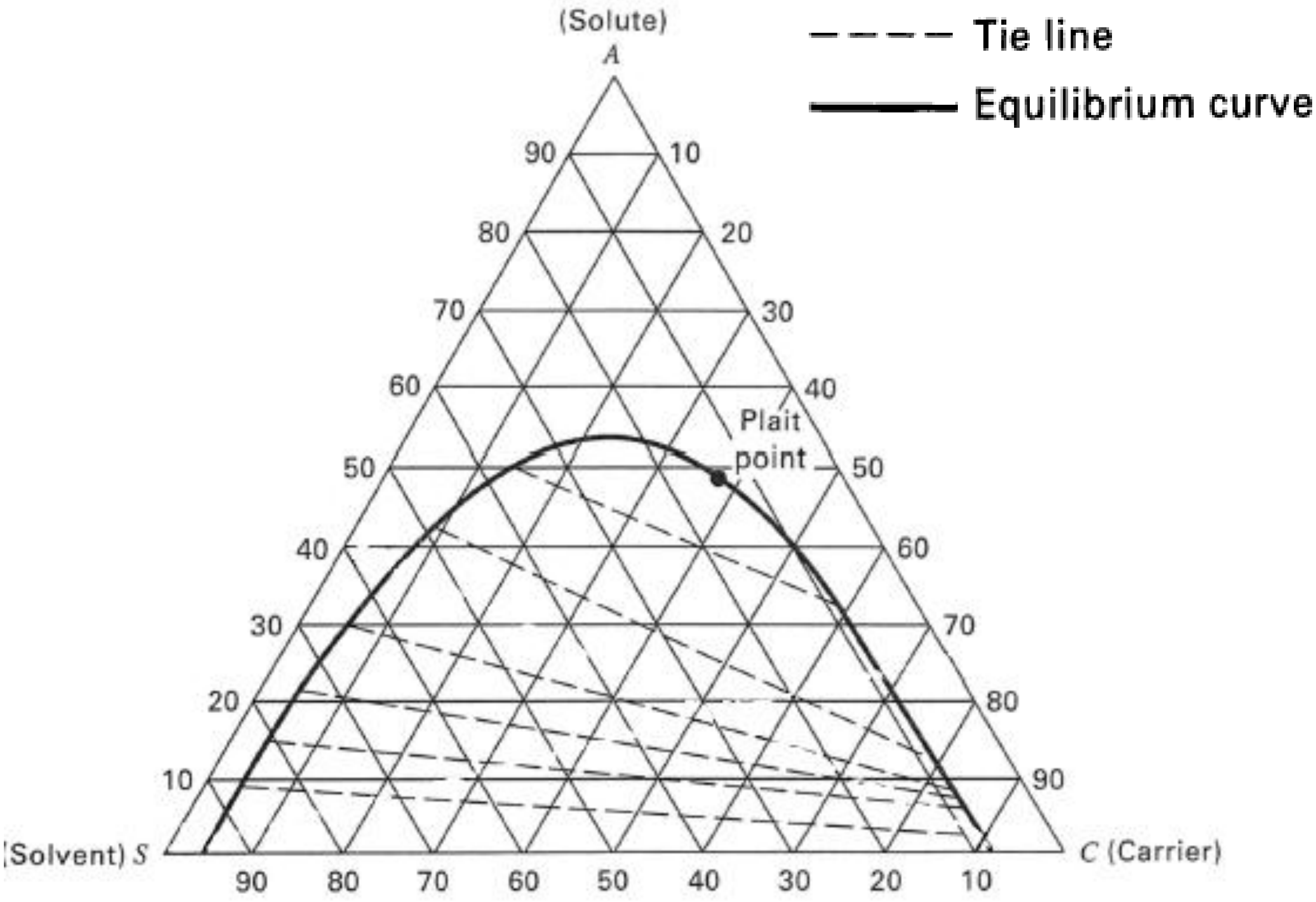
b) Mass balances

2) Find the amounts and compositions of raffinate and extract phases.

The phase diagram is given on the next slide.

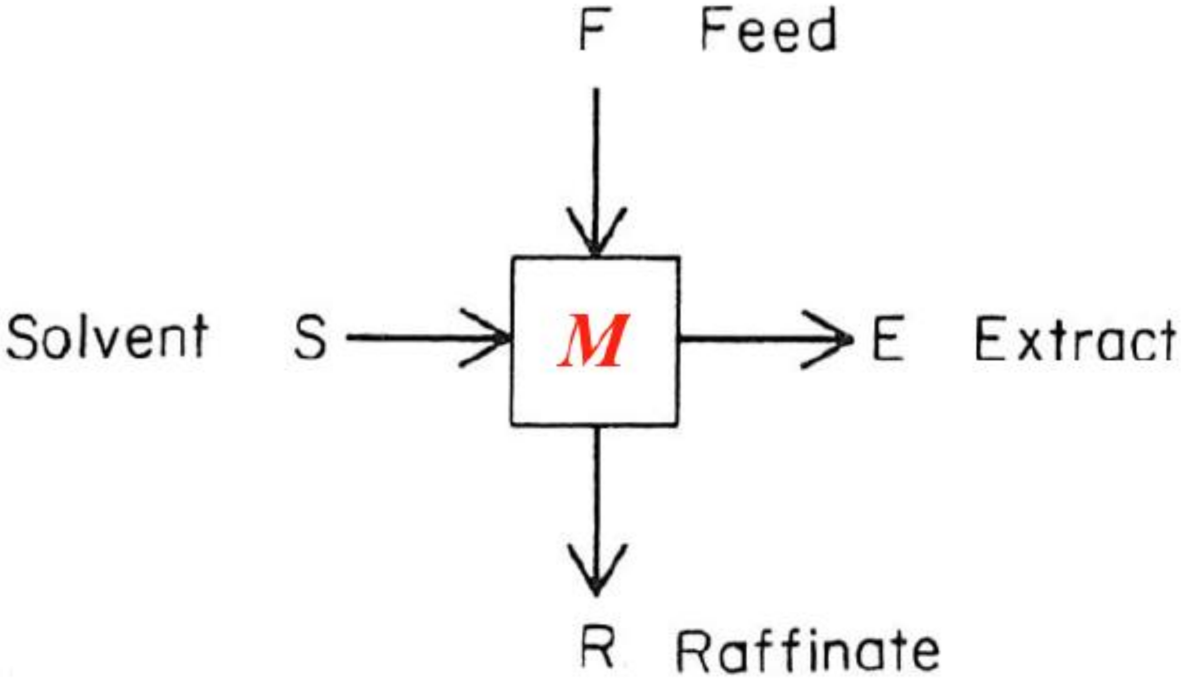
One stage extraction calculations

Example.



One stage extraction calculations

Solution



Feed	Solvent
$F = 250$ kg	$S = 100$ kg
$x_{F,A} = 0.24$	$x_{S,A} = 0.0$
$x_{F,C} = 0.76$	$x_{S,C} = 0.0$
$x_{F,S} = 0.00$	$x_{S,S} = 1.0$

One stage extraction calculations

Solution

Overall composition of mixture using phase diagram

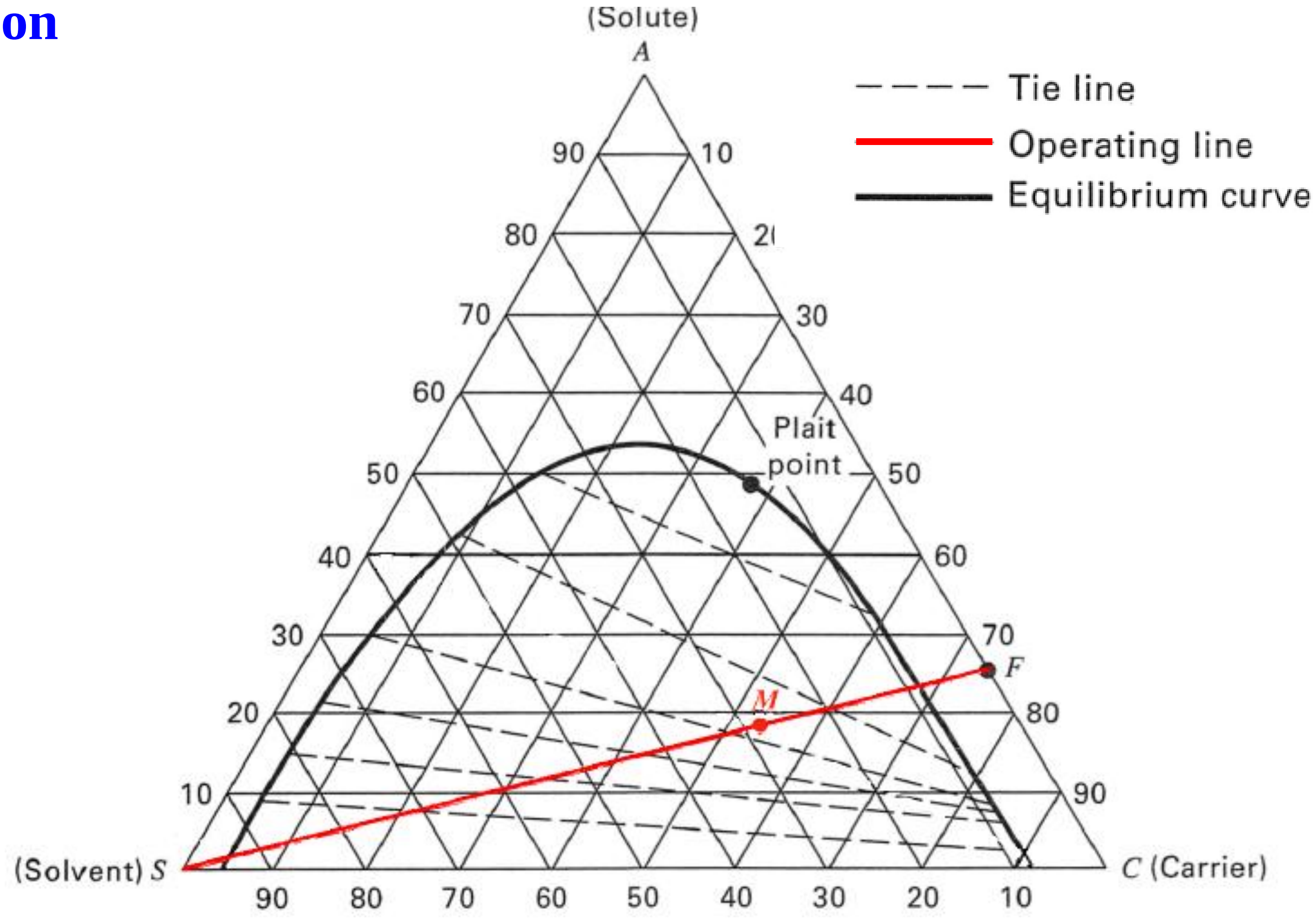
- Locate the feed point (F) on the phase diagram
- Locate the solvent point (S) on the phase diagram
- Connect them by straight line (**operating line**).
- The mixing point must be on operating line, locate it:
  - Apply total mass balance: $\boxed{M = S + F = 100 + 250 = 350\,\text{kg}}$
  - Measure the distance $\overline{SF}$ using the ruler
  - Apply lever-arm rule:

$$\boxed{\frac{S}{M} = \frac{100}{350} = 0.2857 = \frac{\overline{MF}}{\overline{SF}}}$$

  - get $\overline{MF}$ to locate M.
- Read the mixture composition

One stage extraction calculations

Solution



Answer: $x_{M,A} = 0.17; \ x_{M,C} = 0.54; \ x_{M,S} = 0.29$

One stage extraction calculations

Solution

Overall composition using mass balances:

Feed	**Solvent**
$F = 250\text{ kg}$	$S = 100\text{ kg}$
$(x_A)_F = 0.24$	$(x_A)_S = 0.00$
$(x_C)_F = 0.76$	$(x_C)_S = 0.00$
$(x_S)_F = 0.00$	$(x_S)_S = 1.00$

$$M = F + S = 250 + 100 = 350\text{ kg}$$

$$\begin{aligned}(x_A)_M M &= (x_A)_F F + (x_A)_S S \\ &= 0.24(250) + 0(100) = 60\text{ kg} \\ (x_A)_M &= 60/350 = 0.171\end{aligned}$$

$$\begin{aligned}(x_C)_M M &= (x_C)_F F + (x_C)_S S \\ &= 0.76(250) + 0(100) = 190\text{ kg} \\ (x_C)_M &= 190/350 = 0.543\end{aligned}$$

$$\begin{aligned}(x_S)_M M &= (x_S)_F F + (x_S)_S S \\ &= 0(250) + 1(100) = 100\text{ kg} \\ (x_S)_M &= 100/350 = 0.286\end{aligned}$$

One stage extraction calculations

Composition of raffinate and extract phases :

- Interpolate the tie line that passes through the mixture point M.
- The read the composition of R and E phases.

$$x_{R,A} = 0.10$$

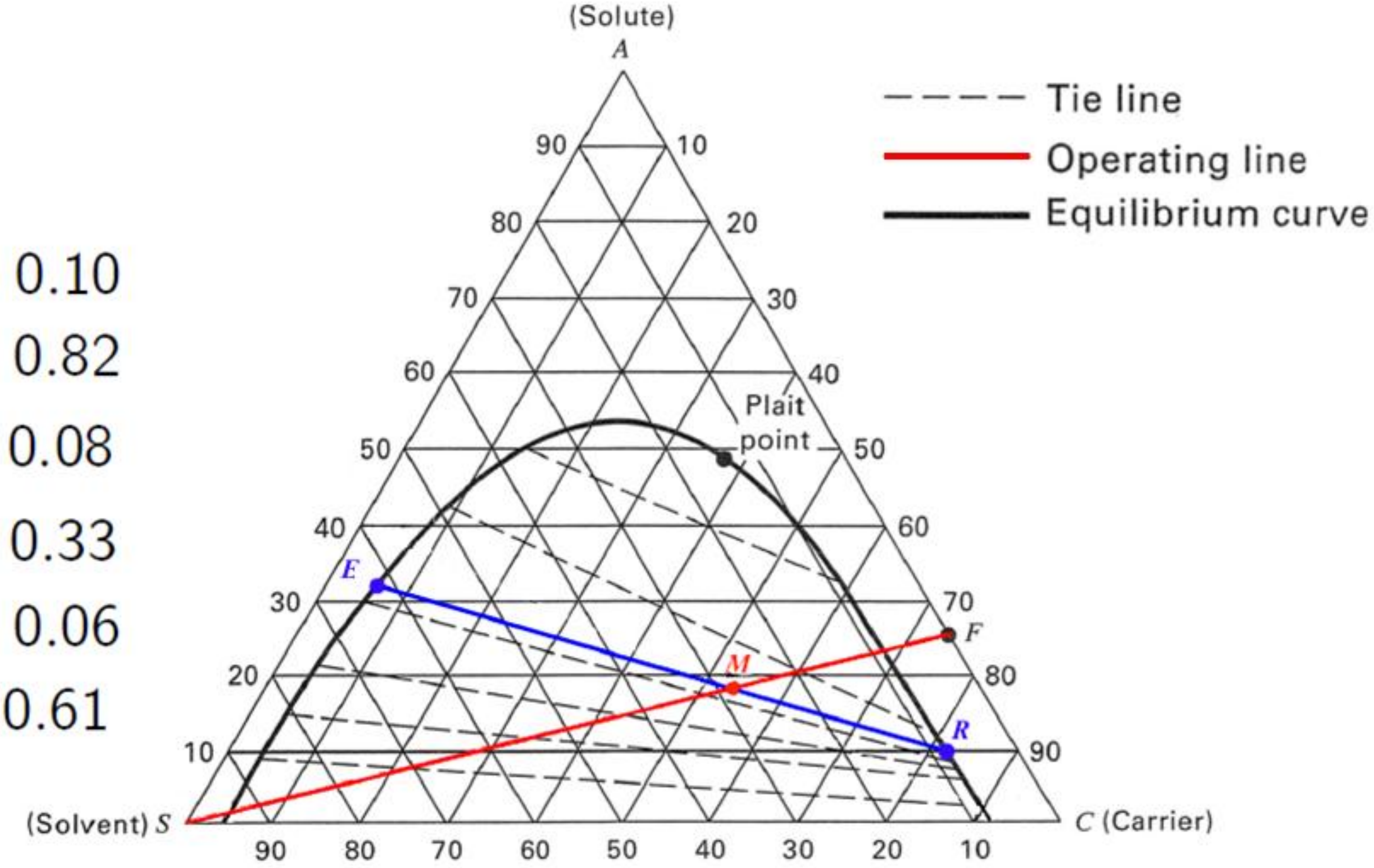
$$x_{R,C} = 0.82$$

$$x_{R,S} = 0.08$$

$$y_{E,A} = 0.33$$

$$y_{E,C} = 0.06$$

$$y_{E,S} = 0.61$$



One stage extraction calculations

Solution

- To find the amount of raffinate (R) and extract (E):

  - Apply lever rule:

$$\boxed{\frac{R}{M}=\frac{R}{350}=\frac{\overline{ME}}{\overline{RE}}}$$

  - Measure the distances $\overline{RE}$ and $\overline{ME}$ using the ruler.

- Calculate R
- Apply overall mass balance to calculate E:

$$\boxed{M=S+F=100+250=350\,\text{kg}=\text{E}+\text{R}}$$

$$\boxed{\text{Answer: E}=128\,\text{kg}\;;\;\;\text{R}=222\;\text{kg}}$$

Solute recovery for one stage extraction

- **Recovery**≡ fraction of solute recovered in the extract

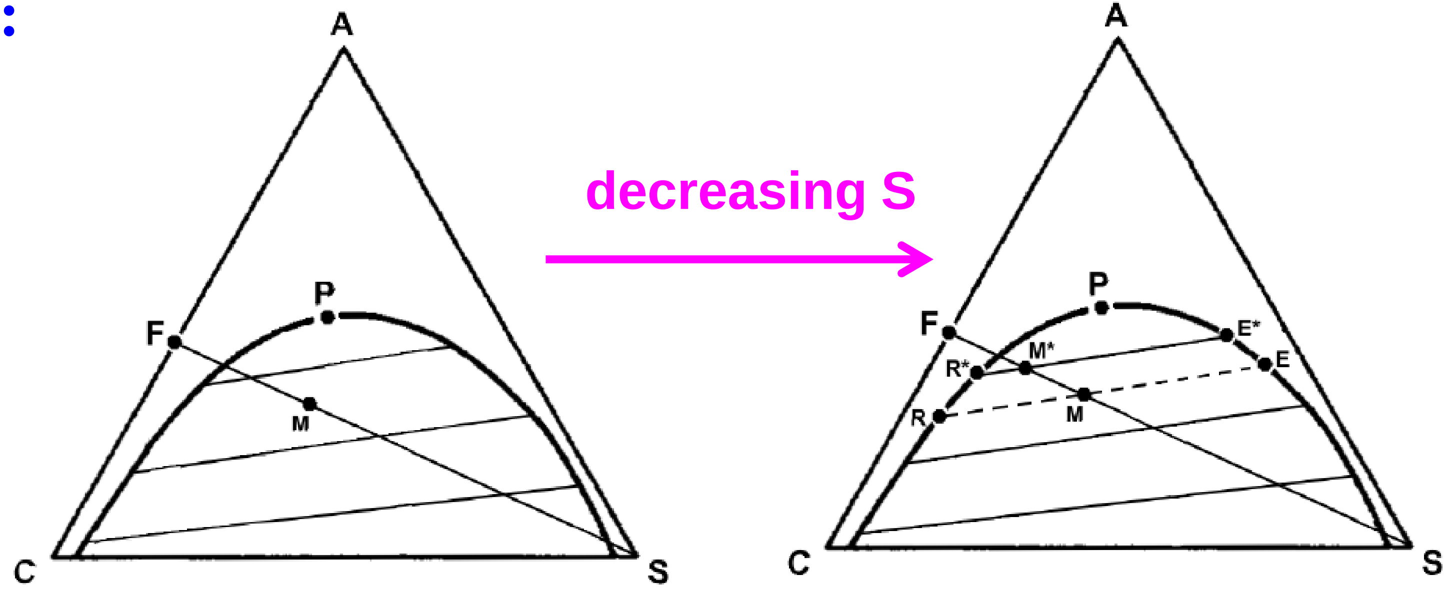
$$\boxed{\text{Solute recoverey}=\frac{Fx_{F,A}-Rx_{R,A}}{Fx_{F,A}}=1-\frac{Rx_{R,A}}{Fx_{F,A}}}$$

Example. Find the recovery in the previous example

$$\text{Solute recoverey}=1-\frac{(222)(0.1)}{(250)(0.24)}=63\%$$

Effect of solvent flow rate (S)

Illustration:

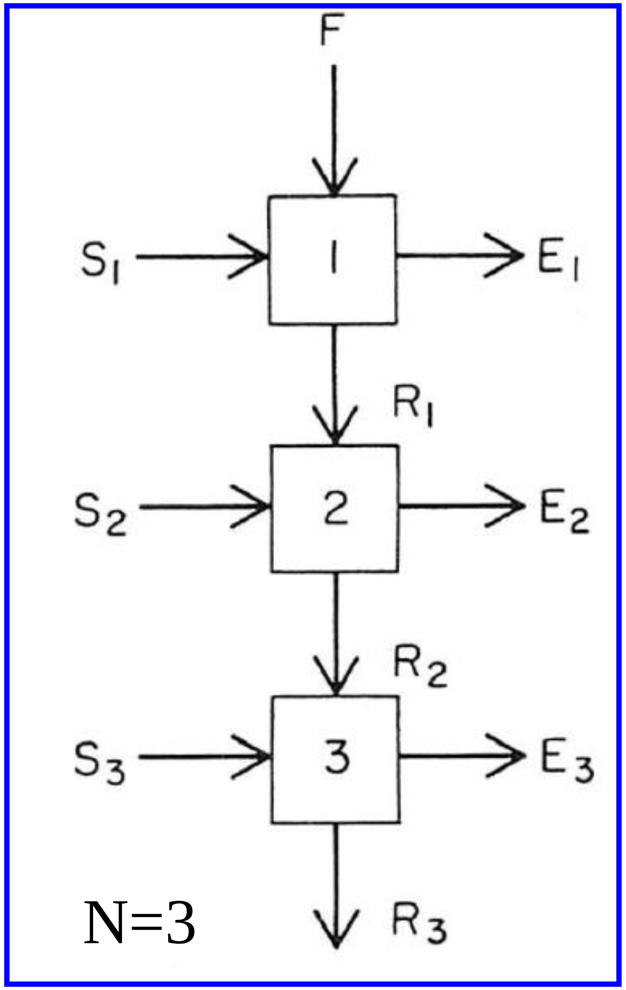


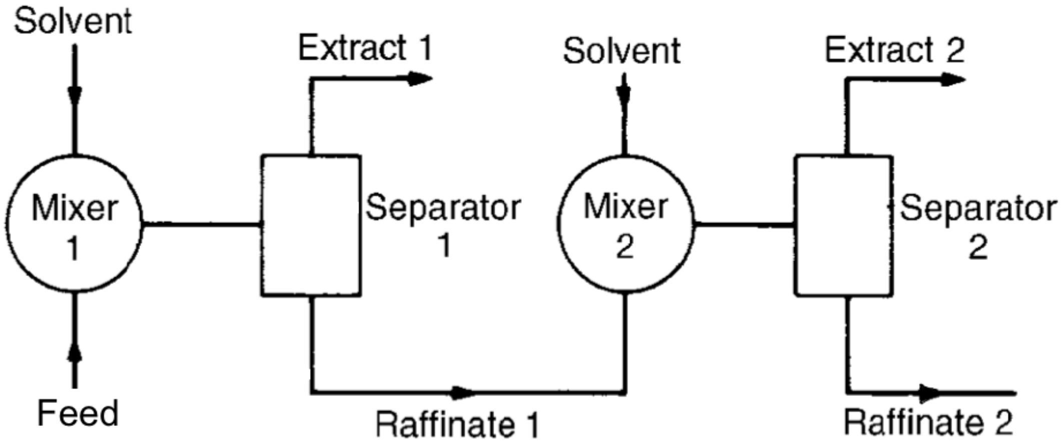
- Based on the above phase diagram, as S decreases the mixture point M on the operating line will move toward the feed point F to be at new location M*. This M* has new equilibrium tie line at which:

$x_{E^*,A} > x_{E,A}$ **(Solute concentration increases with S decrease)**

$x_{R^*,A} > x_{R,A}$ **(Solute recovery decreases with S decrease)**

Cross-current stage extraction





- **Recovery**≡ fraction of solute recovered

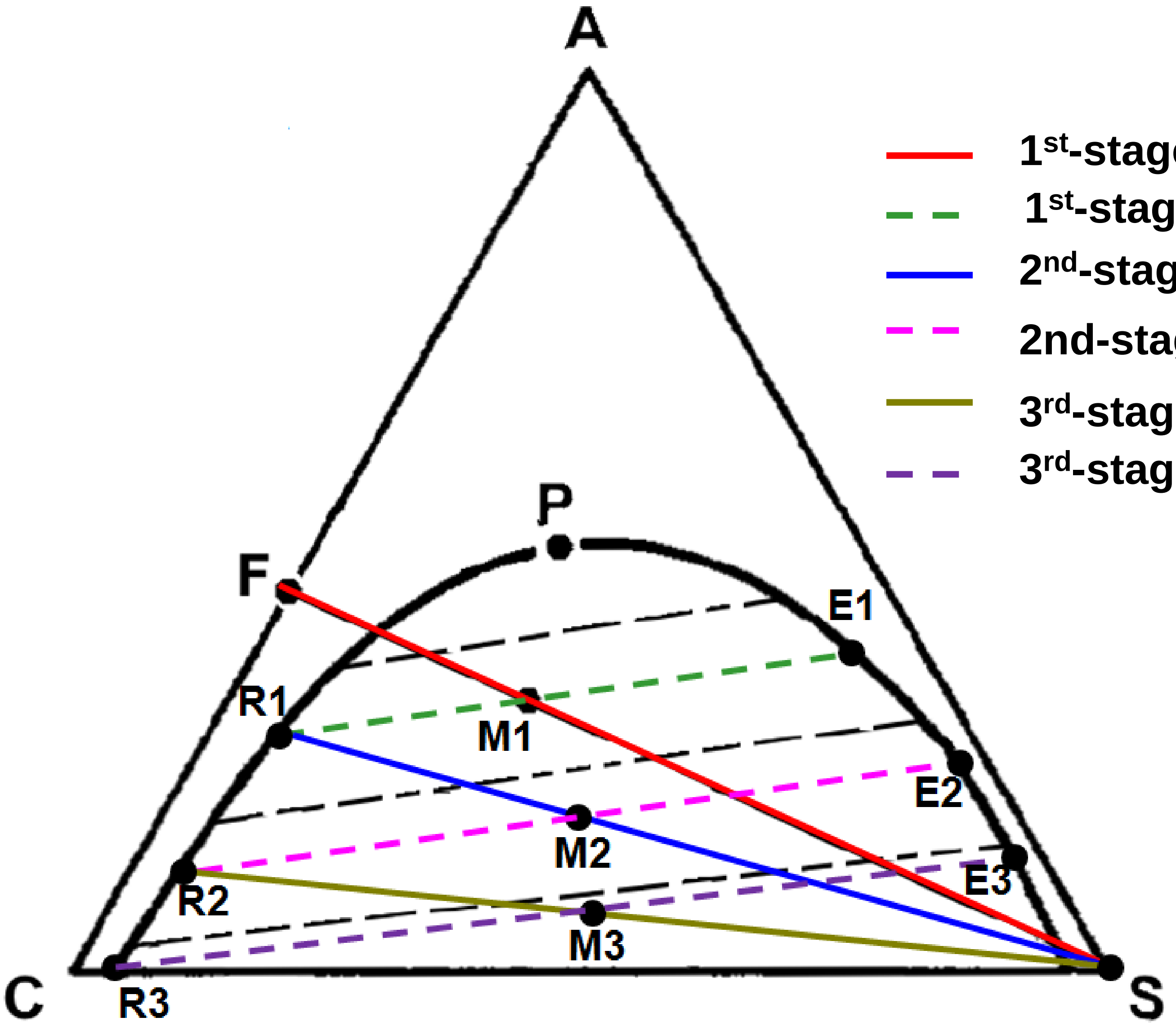
$$\boxed{\text{Solute recovery} = 1 - \frac{R_N x_{R_N,A}}{F x_{F,A}}}$$

N: number of stages

- Overall **solute concentration** in the extract

$$\boxed{\bar{y}_{E,A} = \sum_{i=1}^{N} E_i y_{E_i,A} \Big/ \sum_{i=1}^{N} E_i}$$

Cross-current stage extraction



— 1st-stage operating line

- - 1st-stage tie line

— 2nd-stage operating line

- - 2nd-stage tie line

— 3rd-stage operating line

- - 3rd-stage tie line

Cross-current stage extraction

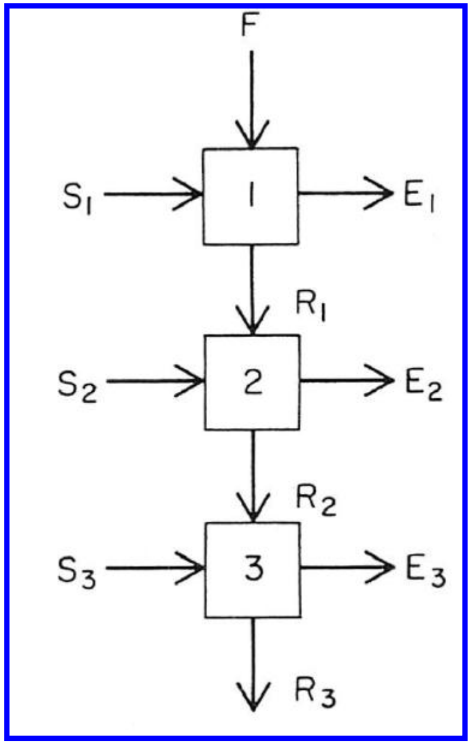
- **Important questions arise:**

- What solvent flow rate?
- How solvent flow rates are distributed among stages?
- How many stages are required to…?

Trade off: solute recovery and overall solute concentration from the extract

Cross-current stage extraction

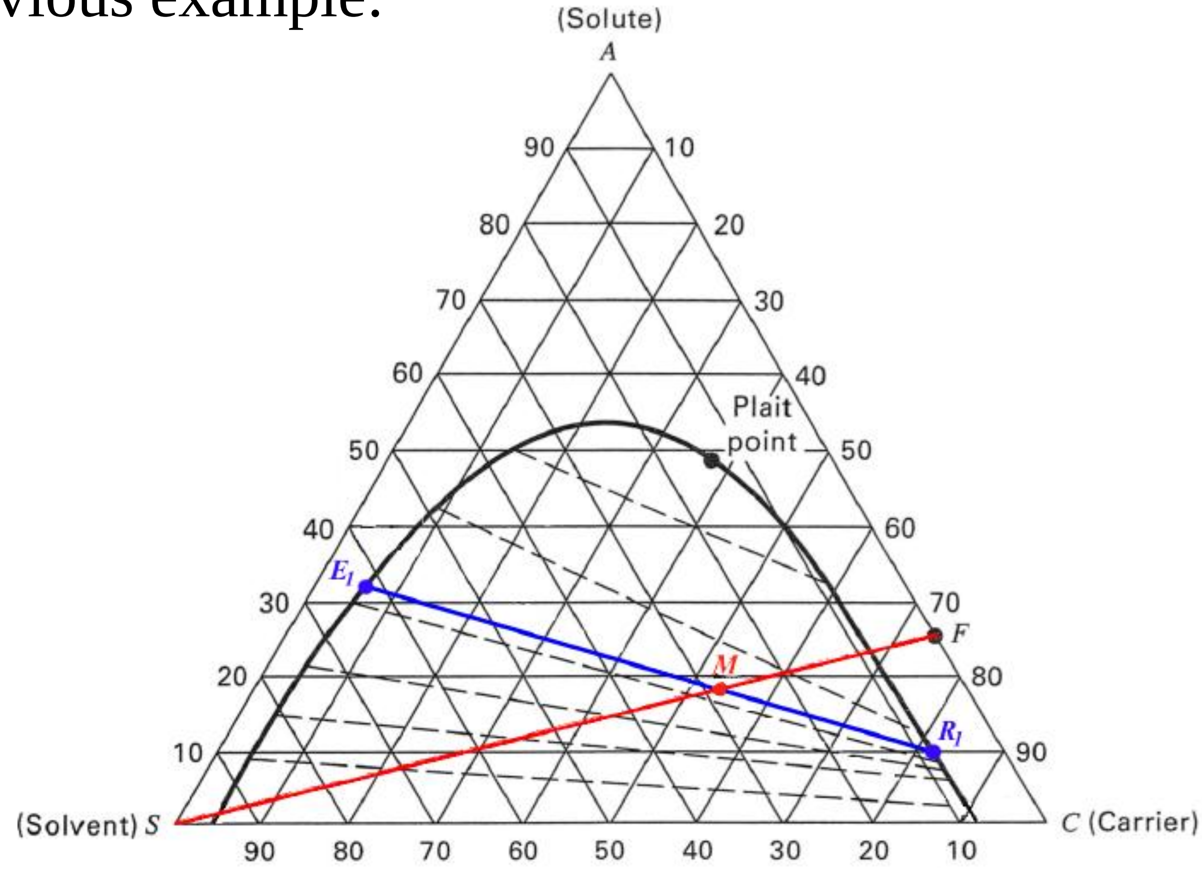
Example. Cross-current mixer-settler extraction units are used for extraction process of 250 kg of feed which contains 24 wt% solute(A) and 76 wt% carrier(C). Each stage is supplied with pure 100 kg solvent (S). Find the minimum number of stages required to achieve at least 85% solute recovery. Find the corresponding overall solute concentration in the extract.



Cross-current stage extraction

Solution:

First stage from previous example:



$$R_1 = 222\text{kg}; \quad x_{R_1,A} = 0.10; \quad x_{R_1,C} = 0.82; \quad x_{R_1,S} = 0.08$$

$$E_1 = 128\text{kg}; \quad y_{E_1,A} = 0.33; \quad y_{E_1,C} = 0.06; \quad y_{E_1,S} = 0.61$$

Cross-current stage extraction

Solution:

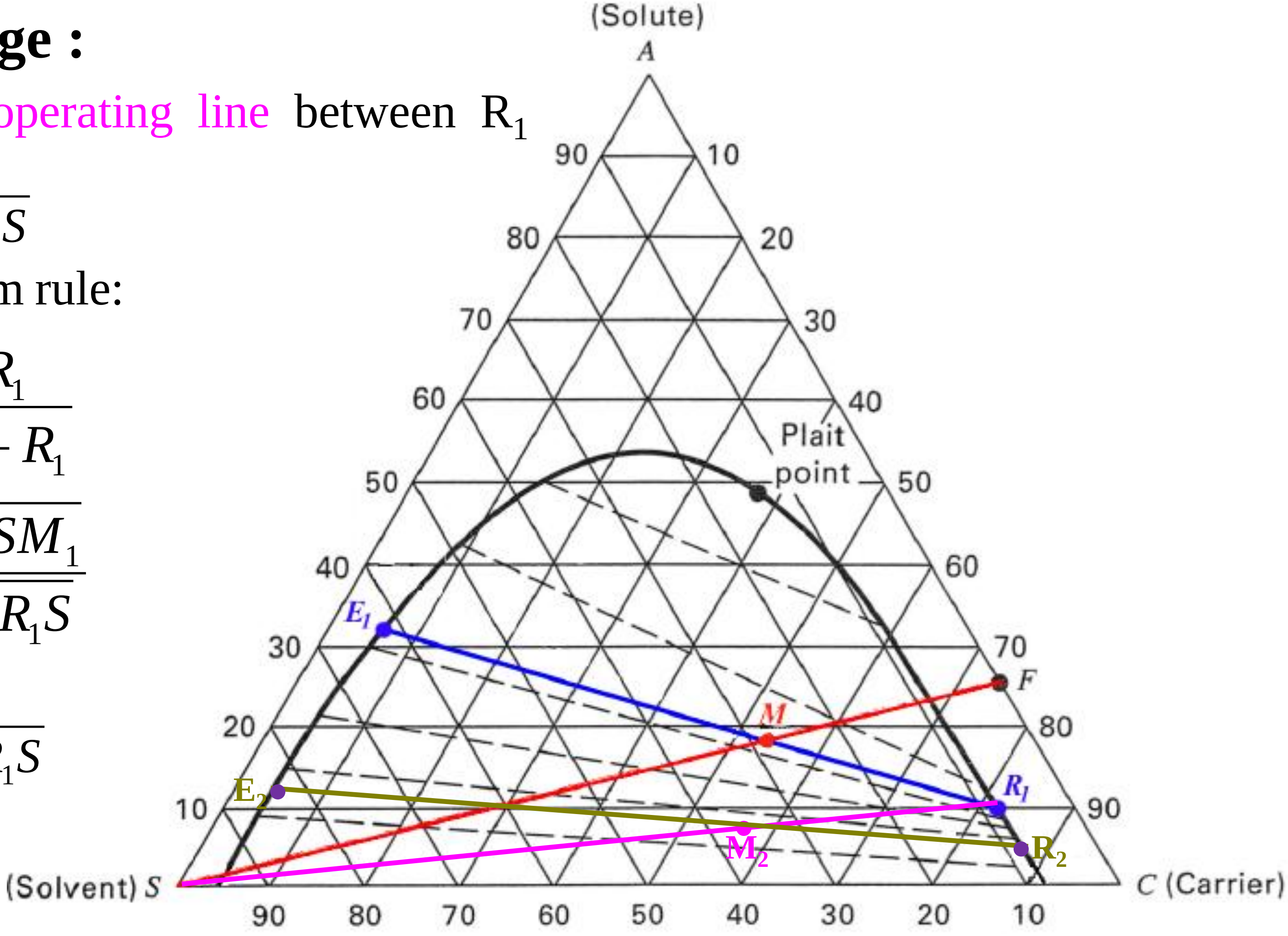
Second stage :

- Draw the operating line between R_1 and S.
- Measure $\overline{R_1S}$
- Use liver-arm rule:

$$\frac{R_1}{M_2} = \frac{R_1}{S + R_1}$$

$$= \frac{222}{322} = \frac{\overline{SM_1}}{\overline{R_1S}}$$

- Calculate: $\overline{R_1S}$
- Locate M_2



Cross-current stage extraction

Solution:

Second stage :

- Interpolate the tie line for stage 2
- Read composition in R_2 and E_2: $x_{R_2,A}=0.04; x_{R_2,C}=0.88; x_{R_2,S}=0.08$

$$y_{E_2,A}=0.11; y_{E_2,C}=0.05; y_{E_2,S}=0.84$$

- Measure $\overline{E_2R_2}$ and $\overline{E_2M_2}$
- Apply lever arm rule to get R_2: $\frac{R_2}{M_2}=\frac{R_2}{S+R_1}=\frac{R_2}{322}=\frac{\overline{E_2M_2}}{\overline{E_2R_2}} \rightarrow R_2=200\,\text{kg}$

$$E_2=M_2-\mathrm{R}_2=122\,\text{kg}$$

- Solute recoverey $=1-\frac{R_2x_{R_2,A}}{Fx_{F,A}}=1-\frac{(200)(0.04)}{(250)(0.24)}=87\%>0.85(\textit{required})$

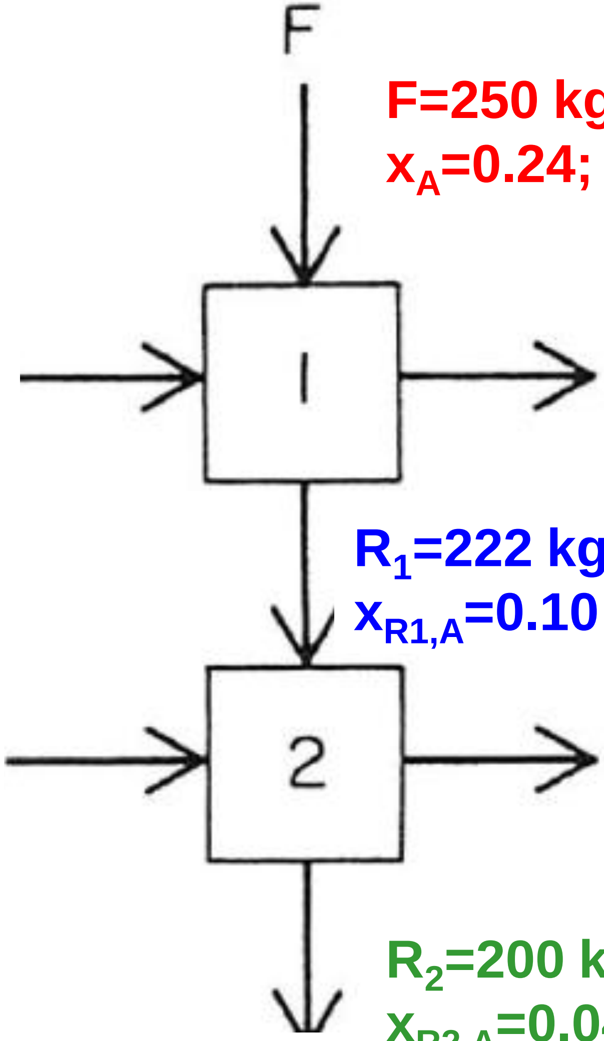
Thus , **two stages** is sufficient to achieve the required solute recovery.

- Overall solute concentration:

$$\bar{\mathrm{y}}_{\mathrm{E,A}}=\sum_{i=1}^{2}E_iy_{E_i,A}\Big/\sum_{i=1}^{N}E_i=\left(E_1y_{E_1,A}+E_2y_{E_2,A}\right)/\left(E_1+E_2\right)=0.22$$

Cross-current stage extraction

Use mass balances on the overall system and on each stage to verify the ans



F=250 kg
x_A=0.24; x_C=0.76

S_1=100 kg

E_1=128 kg
$y_{E1,A}$=0.33; $y_{E1,C}$=0.06

R_1=222 kg/h
$x_{R1,A}$=0.10; $X_{R1,C}$=0.82

S_2=100 kg

E_2=122 kg
$y_{E2,A}$=0.11; $y_{E2,C}$=0.05

R_2=200 kg
$x_{R2,A}$=0.04; $X_{R2,C}$=0.88

Cross-current stage extraction

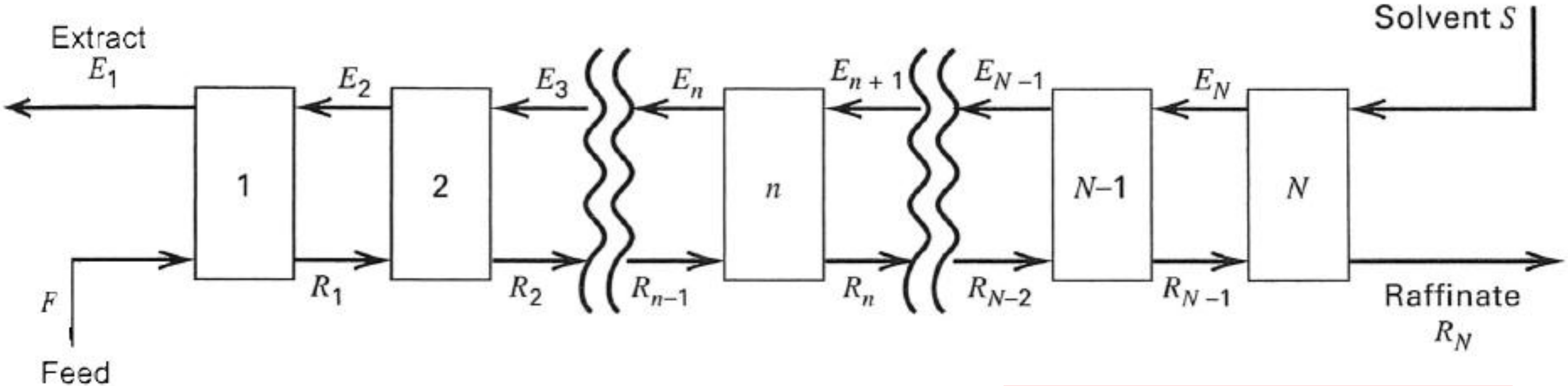
Exercise . Consider a system for which you have been given the ternary diagram . A = solute, S = solvent (100% pure), C = carrier. For the overall two-stage cross-current system, the feed, F enters at 112 kg/hr with composition of 25 wt% solute and 75 wt% carrier. Each stage is provided with pure solvent of 50 kg/h. Find:

a. Overall recovery [**answer: 93%**].

b. Overall concentration of combined extract streams [**answer: 21%**]

Counter-current stage extraction

N units in counter-current arrangement:



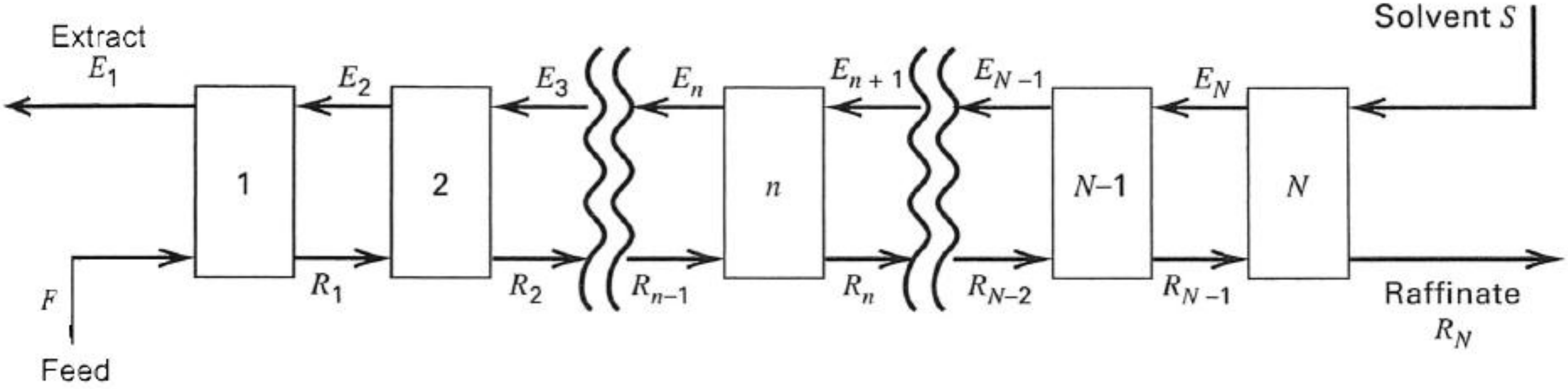
n: index for stage number

- "**Re-use**" of solvent
- E_n and R_n for n=1,…N leaving each stage are in equilibrium [they are determined via the **tie line**]

- **Solute Recovery:** $$\text{Solute recovery} = 1 - \frac{R_N x_{R_N,A}}{F x_{F,A}}$$

- **Overall solute concentration:** $\bar{y}_{E,A} = y_{E_1,A}$

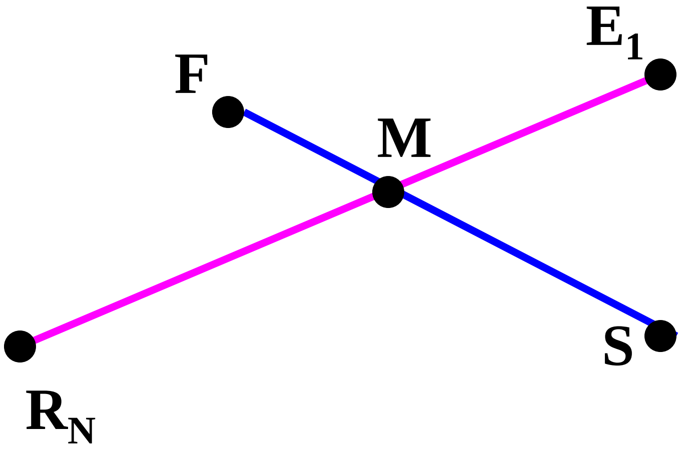
Counter-current stage extraction



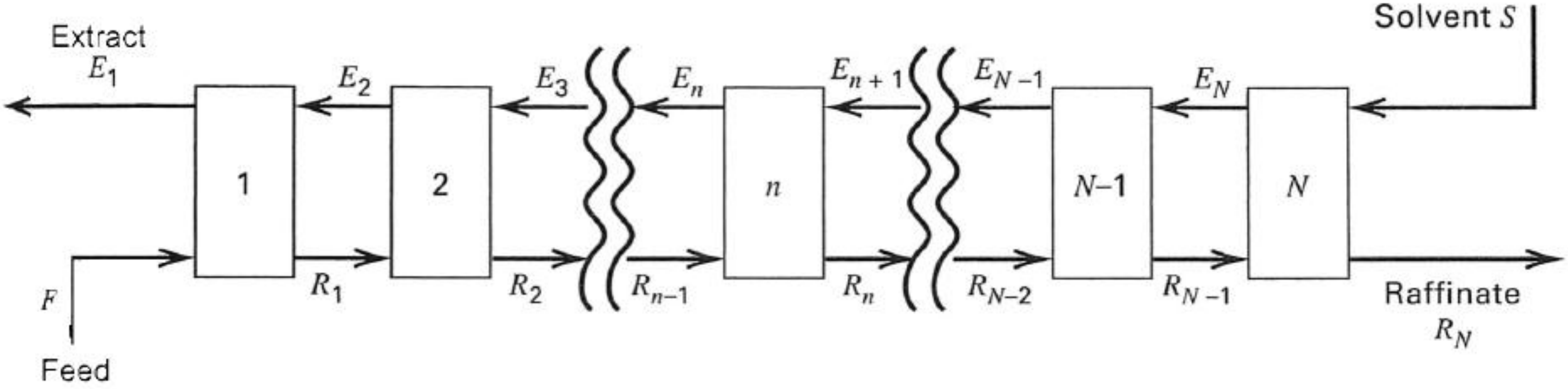
Apply st. st. total mass balance on the overall system:

$\boxed{F + S = E_1 + R_N = M}$ →Lever-arm rule:

- F, S, and M must located on the same straight line and mixture point M is between F and S.
- E_1, R_N, and M must located on the same straight line and mixture point M is between E_1 and R_N.



Counter-current stage extraction



Apply total mass balance on each stage:

First stage: $\boxed{\mathrm{F} + E_2 = E_1 + R_1 \ \ or\ \ \mathrm{F} - E_1 = R_1 - E_2}$

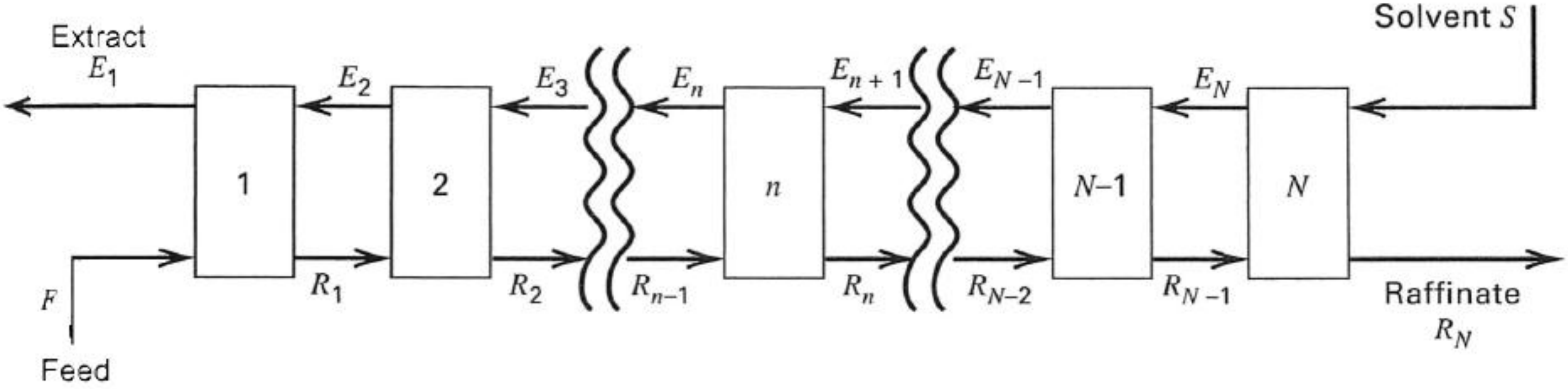
Second stage: $\boxed{E_2 + R_2 = E_3 + R_1 \ ;or\ R_1 - E_2 = R_2 - E_3}$

and in general for stages n = 2,…,N-1

$$\boxed{E_n + R_n = E_{n+1} + R_{n-1} \ ;\ n = 2,..,\ N-1 \ \ or\ R_{n-1} - E_n = R_n - E_{n+1}}$$

Last stage: $\boxed{\mathrm{R}_{\mathrm{N\text{-}1}} + S = E_N + R_N \ \ or\ \ \mathrm{R}_{\mathrm{N\text{-}1}} - E_N = R_N - S}$

Counter-current stage extraction

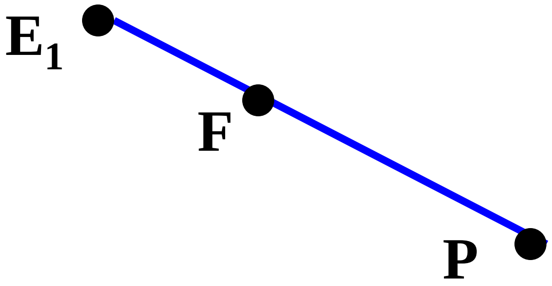


It is clear that

$$\boxed{F - E_1 = R_1 - E_2 = R_2 - E_3 = \dots = R_{N-1} - E_N = R_N - S = Constant = P}$$

This means that each flow difference is equal to P

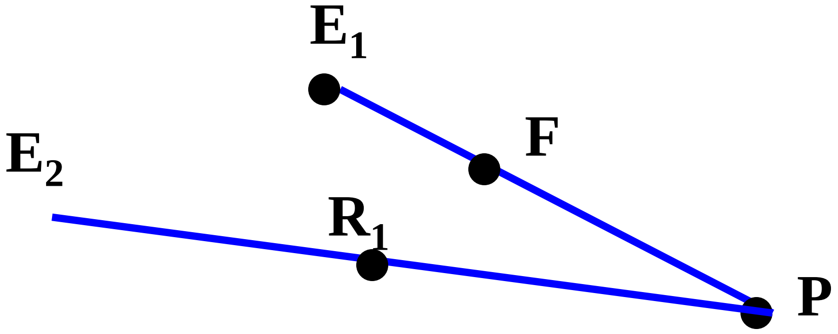
Rearrange as : $\boxed{F = E_1 + P}$ → Lever-arm rule, the three points F, E_1, and P must located on the same straight line and F is between E_1 and P



P: operating point

Counter-current stage extraction

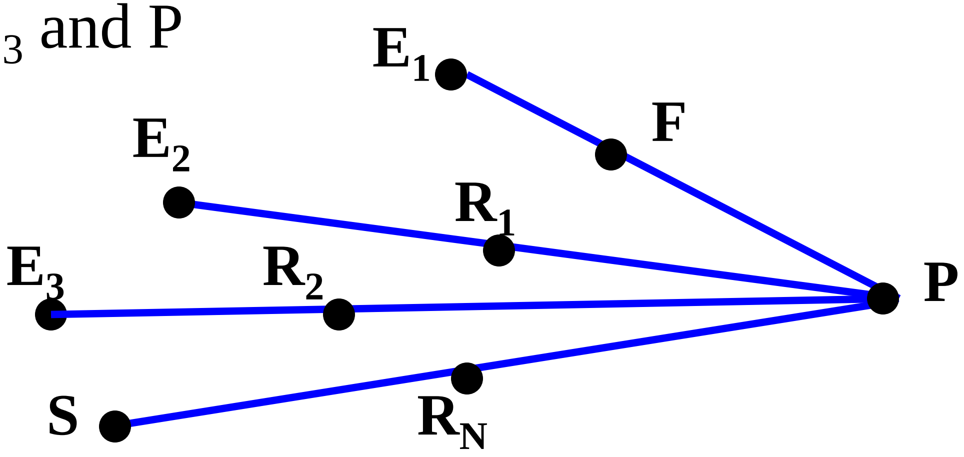
$\boxed{R_1 = E_2 + P}$ → Lever-arm rule, the three points R_1, E_2, and P must located on the same straight line and R_1 is between E_2 and P



$\boxed{R_2 = E_3 + P}$ → Lever-arm rule, the three points R_2, E_3, and P must located on the same straight line and R_2 is between E_3 and P

….and so on to:

$\boxed{R_N = S + P}$



Counter-current graphical solution

General procedure:

1. Connect S and F points with a line.
2. Locate the mixture point M using overall mass balance and lever rule.
3. Either specify E_1 or R_N (we will always know one of them)
4. Connect a straight line through M and passing through the one specified (E_1 or R_N). Cross line with the equilibrium curve give the unspecified one.
5. Locate the operating point P:
   - Connect S through R_N and extrapolate
   - Connect E_1 through F and extrapolate
   - Cross lines at operating point P
6. In general: connect E_n and R_n via equilibrium tie lines

Counter-current graphical solution

Example. Consider a system for which you have been given the ternary diagram . A = solute, S = solvent (100% pure), C = carrier. In a counter-current extraction process, the feed, F enters at 112 kg/hr with composition of 25 wt% solute and 75 wt% carrier. The solvent flow rate is 28 kg/hr.

a. Find the number of theoretical stages required to achieve solute concentration in raffinate of 2.5 wt% (at most).

b. Calculate the overall recovery and solute concentration of the extract stream.

c. Plot solute concentrations in the extract and raffinate streams versus stages number.

$$\boxed{S = 28\,\text{kg/h}\ ; F = 112\ \text{kg/h}\,; x_{R_N,A} = 0.025\ ; x_{F,A} = 0.25;\ x_{F,C} = 0.75}$$

Counter-current graphical solution

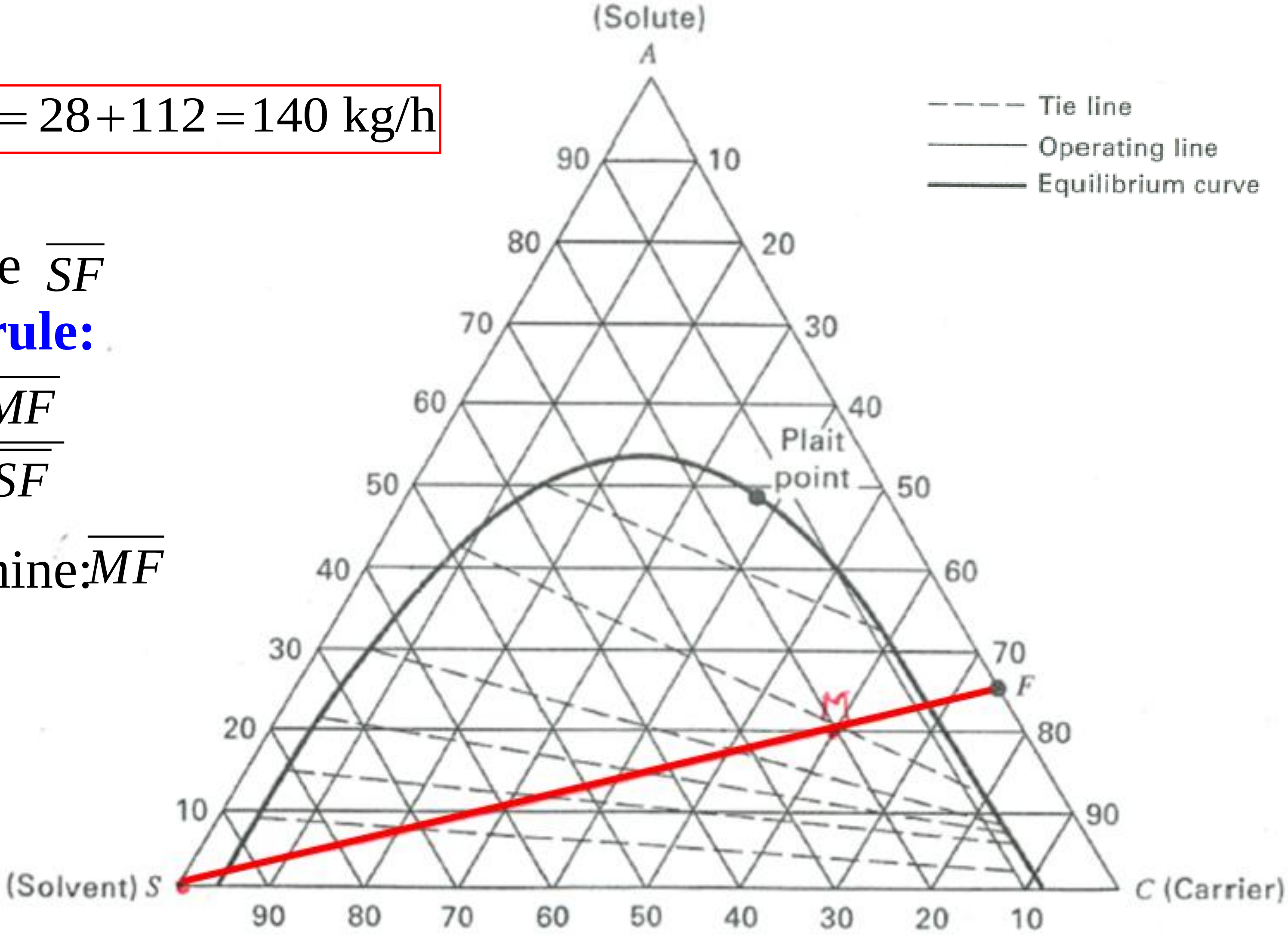
Step 1 and 2: Connect S and F points with a line and locate M.

$$\boxed{M = S + F = 28 + 112 = 140 \text{ kg/h}}$$

- Measure $\overline{SF}$
- **Lever rule:**

$$\frac{S}{M} = \frac{\overline{MF}}{\overline{SF}}$$

- Determine: $\overline{MF}$



Counter-current graphical solution

Step 3 and 4: specify E_1

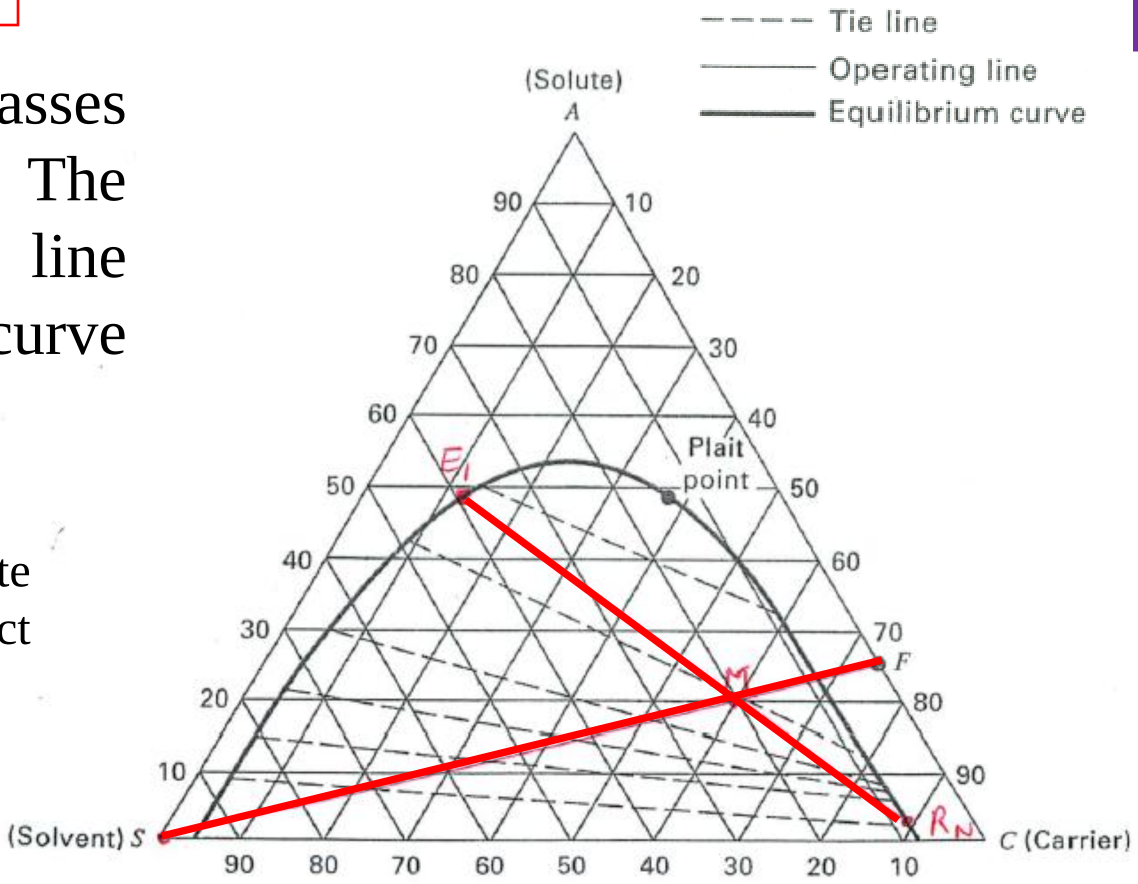
- Locate the point R_N $(x_{R_N,A} = 0.025)$

$$\boxed{M = E_1 + R_N = 140 \text{ kg/h}}$$

- Draw the line passes through M and R_N. The position at which the line crosses the equilibrium curve is E_1
- **Read** $y_{E_1,A} = 48\%$

This is the overall solute concentration of the extract stream:

$$\boxed{\bar{y}_{E,A} = y_{E_1,A} = 48\%}$$



Counter-current graphical solution

Step 3 and 4: specify E_1

- Measure $\overline{E_1R_N}$ and $\overline{ME_1}$
- **Lever rule:**

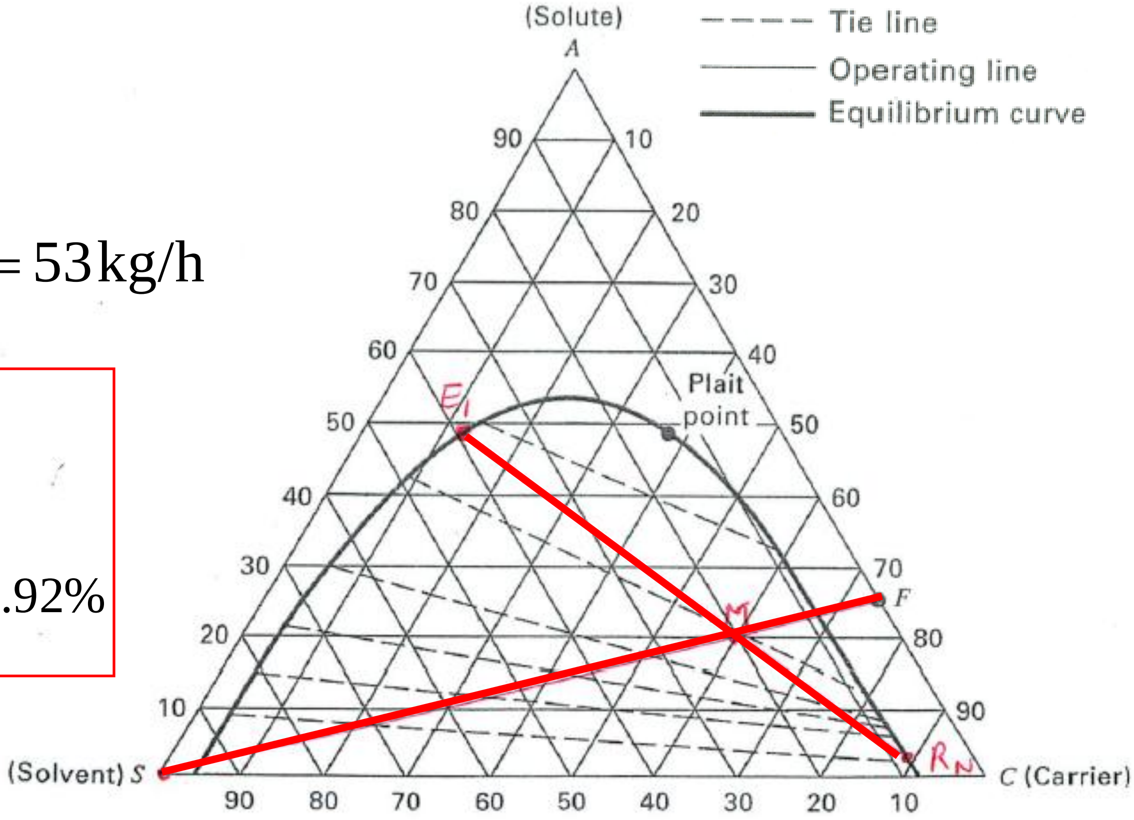
$$\frac{R_N}{M} = \frac{R_N}{140} = \frac{\overline{ME_1}}{\overline{E_1R_N}} = 0.62$$

$$\rightarrow R_N = 87\,\text{kg/h}$$

$$E_1 = M - R_N = 140 - 87 = 53\,\text{kg/h}$$

$$\text{Recovery} = 1 - \frac{R_N x_{R_N,A}}{F x_{F,A}}$$

$$= 1 - \frac{(87)(0.025)}{(112)(0.25)} = 0.92\%$$



Counter-current graphical solution

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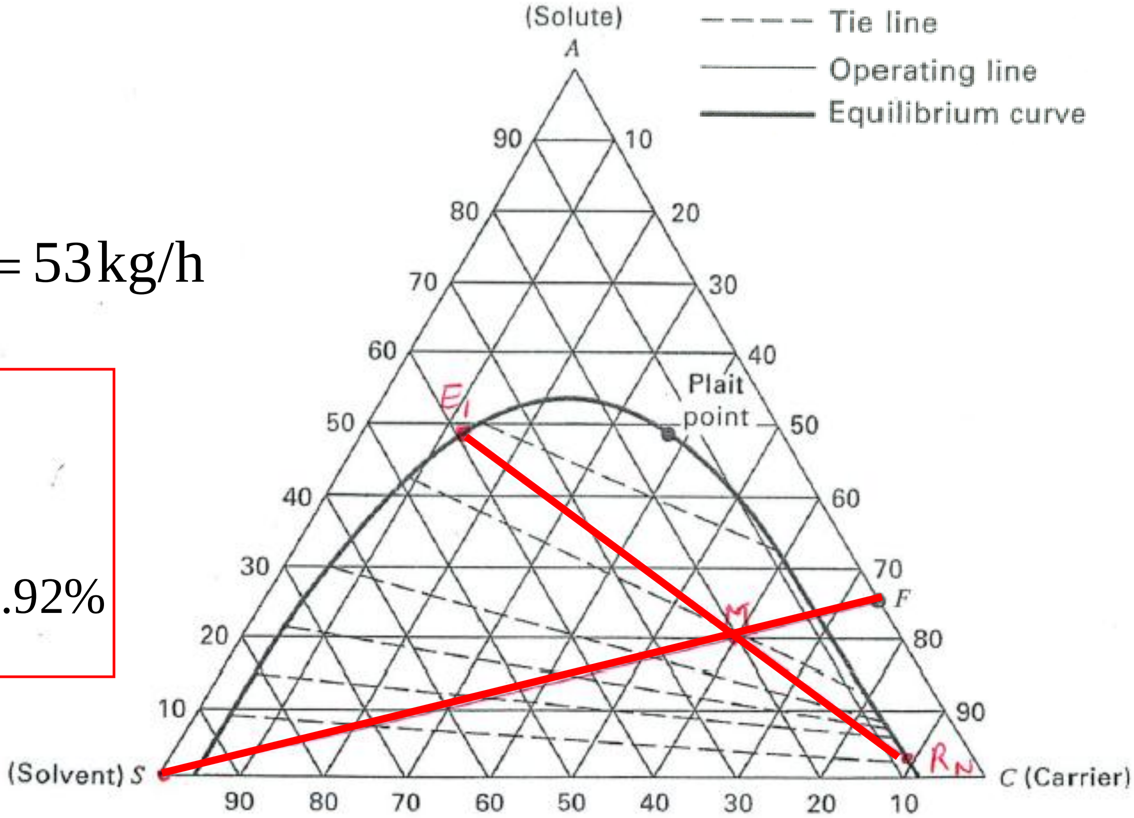
$$\boxed{\begin{aligned}\text{Recovery} &= 1 - \frac{R_N x_{R_N,A}}{F x_{F,A}} \\ &= 1 - \frac{(87)(0.025)}{(112)(0.25)} = 0.92\%\end{aligned}}$$



Counter-current graphical solution

Step 5 : specify the operating point P

$\boxed{F = E_1 + P}$

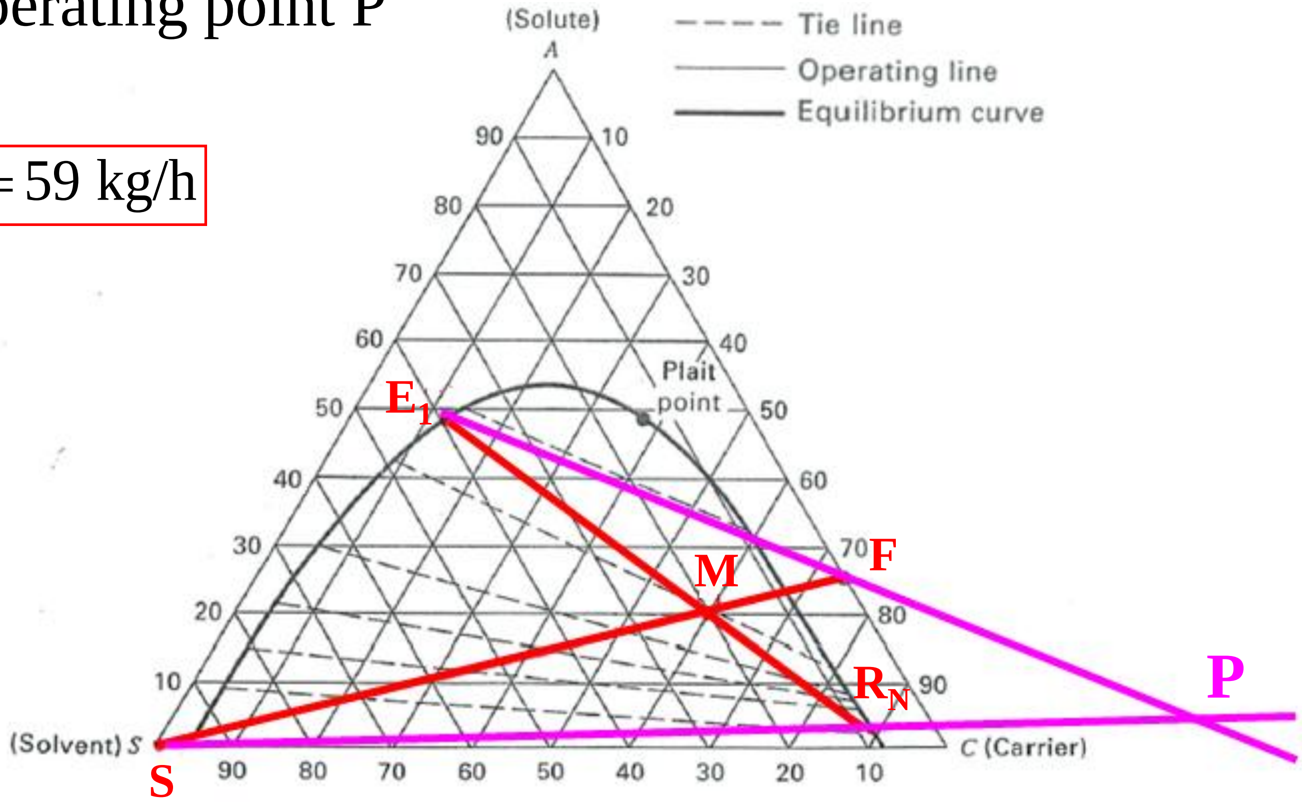
- Connect E_1 through F and extrapolate

$\boxed{R_N = S + P}$

- Connect S through R_N and extrapolate

- Cross lines at operating point P

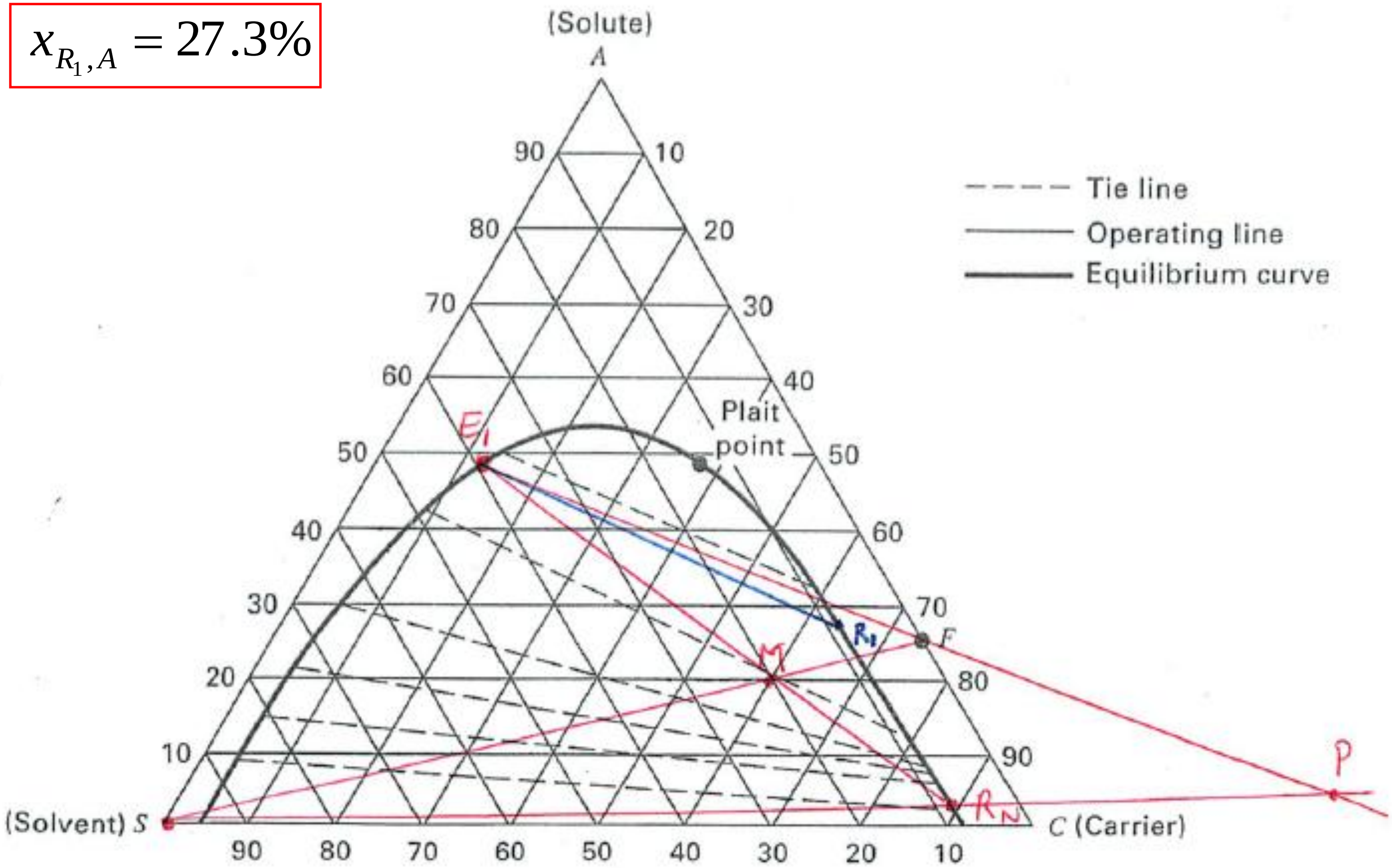
$\boxed{112 = 53 + P \rightarrow P = 59 \text{ kg/h}}$



Counter-current graphical solution

Step 6 : In general: connect E_n and R_n via equilibrium tie lines

- Connect E_1 with R_1 via equilibrium tie lines
- Read: $\boxed{x_{R_1,A} = 27.3\%}$



Counter-current graphical solution

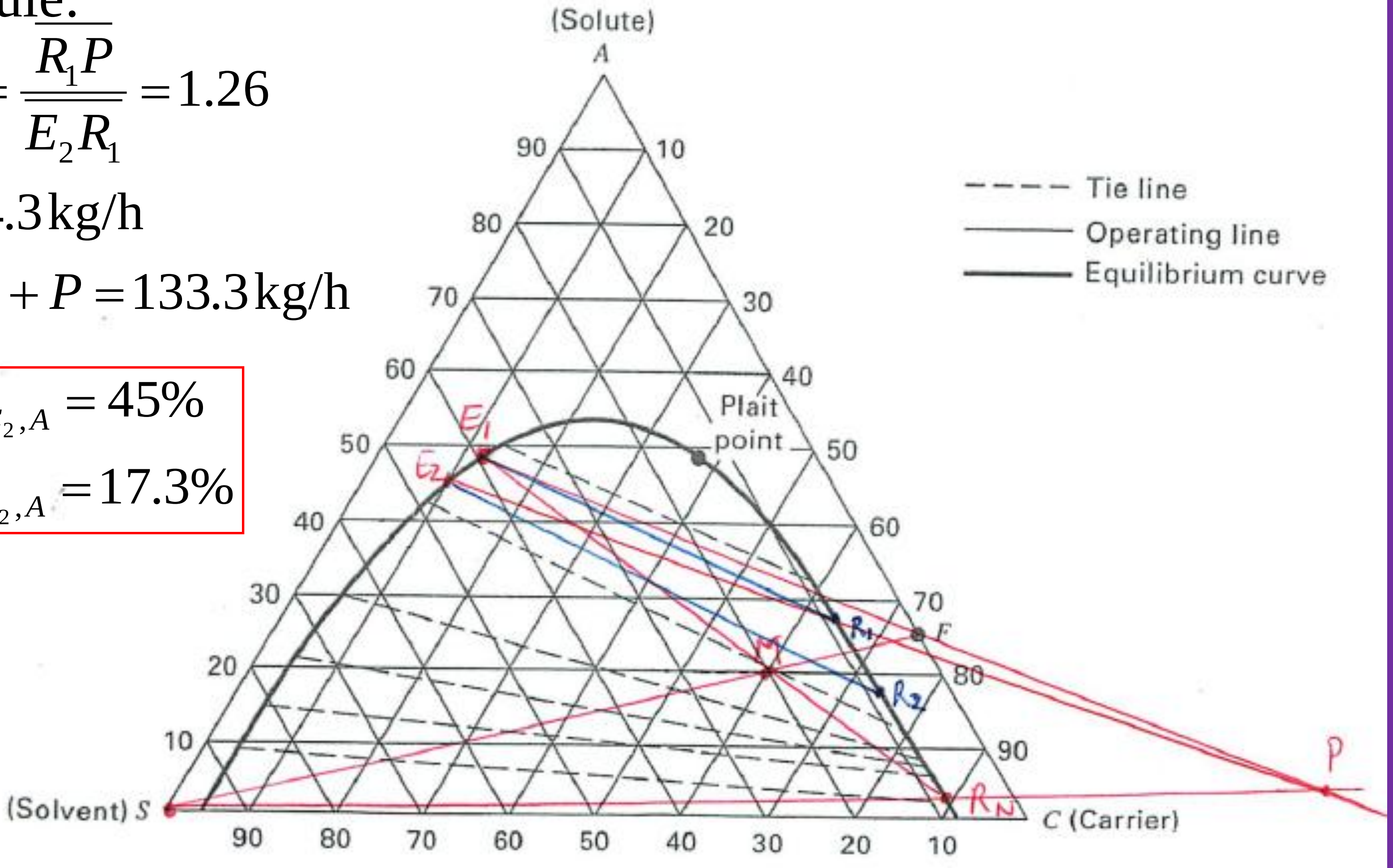
- Connect R_1 with P to extrapolate E_2
- Connect E_2 with R_2 via tie line
- Lever rule:

$$\frac{E_2}{P} = \frac{E_2}{59} = \frac{\overline{R_1P}}{\overline{E_2R_1}} = 1.26$$

$$\rightarrow E_2 = 74.3\,\text{kg/h}$$

$$R_1 = E_2 + P = 133.3\,\text{kg/h}$$

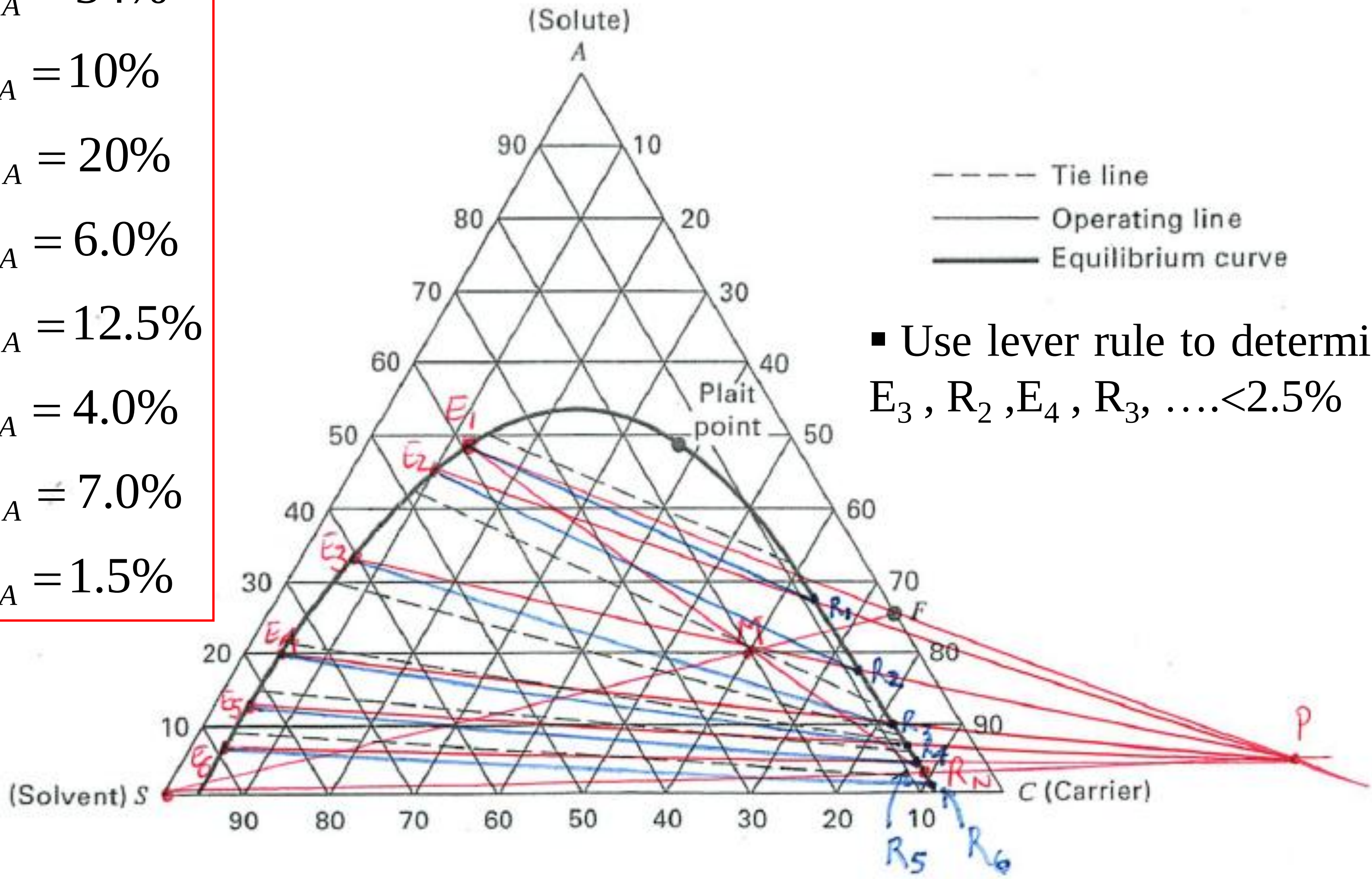
- Read: $\boxed{\begin{array}{l} y_{E_2,A} = 45\% \\ x_{R_2,A} = 17.3\% \end{array}}$



Counter-current graphical solution

- Continue to reach $x_{RN} \leq 2.5\%$

$$y_{E_3,A} = 34\%$$
$$x_{R_3,A} = 10\%$$
$$y_{E_4,A} = 20\%$$
$$x_{R_4,A} = 6.0\%$$
$$y_{E_5,A} = 12.5\%$$
$$x_{R_5,A} = 4.0\%$$
$$y_{E_6,A} = 7.0\%$$
$$x_{R_6,A} = 1.5\%$$

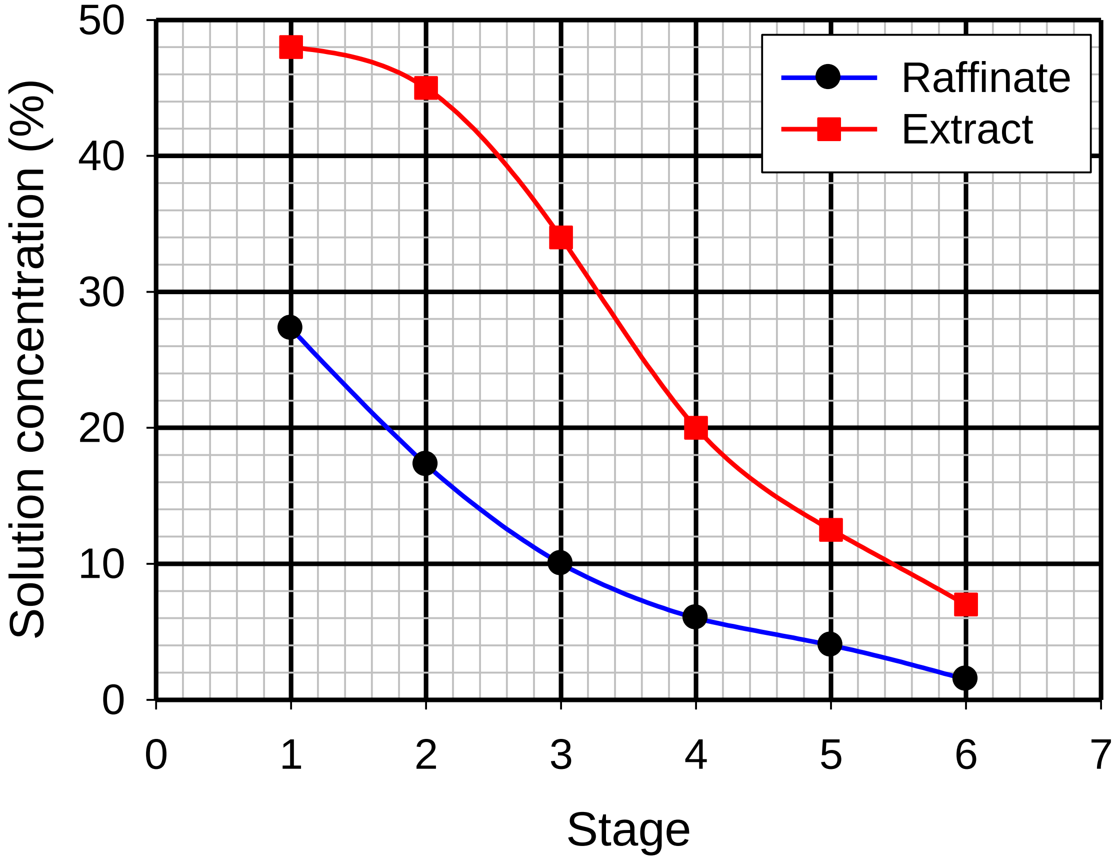


- Use lever rule to determine E_3 , R_2 ,E_4 , R_3, ….<2.5%

Counter-current graphical solution

$$x_{R_6,A} = 1.5\% < 2.5\%$$

N=6 theoretical stages is enough to have solute concentration less than 2.5 wt% in raffinate stream



Counter-current graphical solution

- **Determination of minimum solvent-to-feed ratio (S_{min}/F):**

→ If $x_{RN, A}$ is given :

- Connect F and S with straight line on ternary phase diagram.
- Connect S and R_N with straight line on ternary phase diagram.
- Search for E_1 point on equilibrium curve such that:

  - E_1R_N line intersects with SF line at mixing point with the minimum possible distance about F (maximum possible distance from S. This mixing point is denoted as M_{min}.
  - E_1F line must intersects with SR_N at the operating point denoted as P_{min}.

→ If E_1 is given, similar search approach for R_N can be followed.

Remark. With S_{min}/F, the required number of theoretical stages:

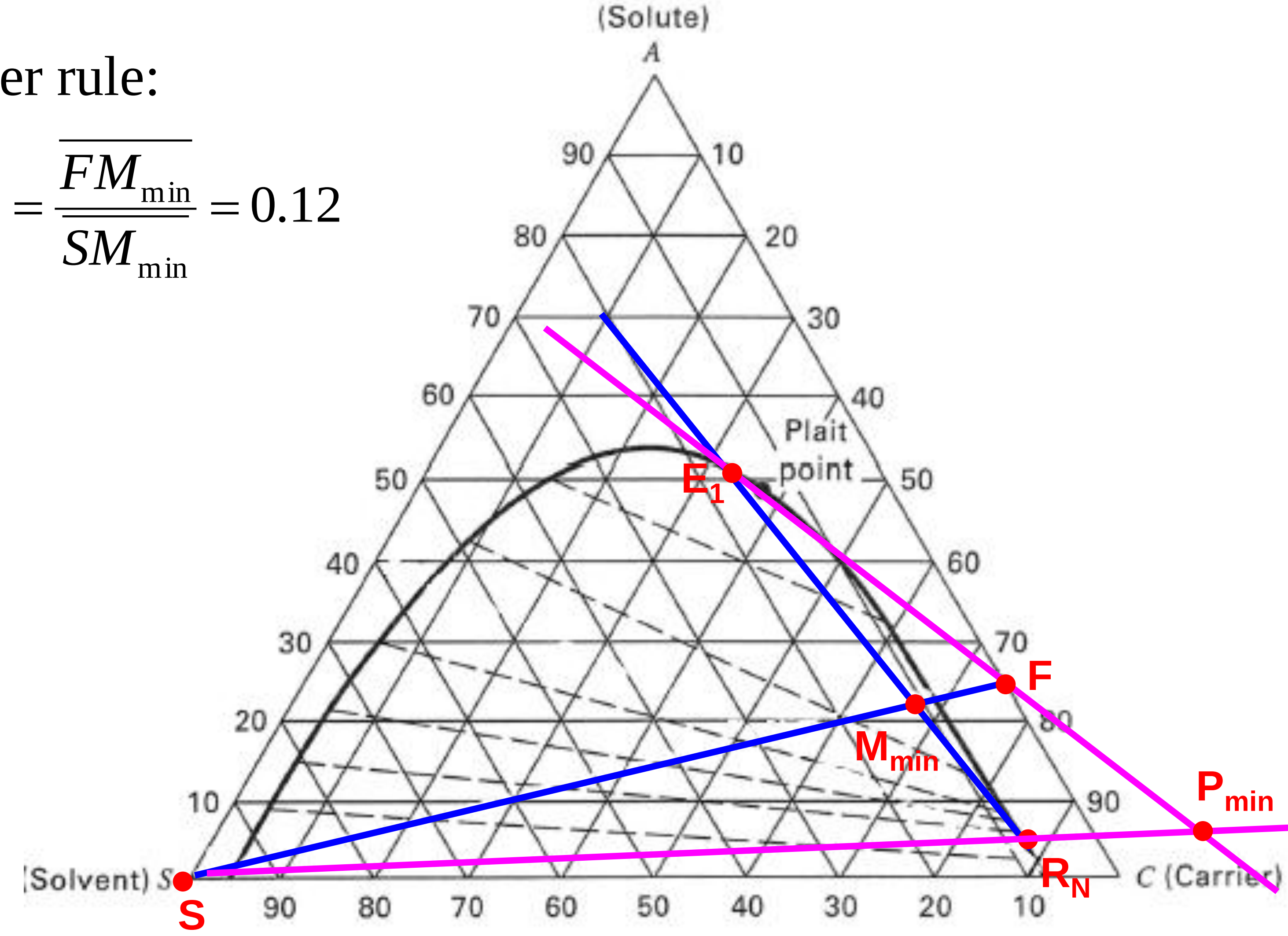
$N \rightarrow \infty$

Counter-current graphical solution

Example. Determine S_{min}/F for the system of previous example

- Lever rule:

$$\frac{S_{\min}}{F} = \frac{\overline{FM_{\min}}}{\overline{SM_{\min}}} = 0.12$$



Counter-current graphical solution

- **Determination of maximum solvent-to-feed ratio (S_{max}/F):**

→ If $x_{RN, A}$ is given :

- Connect F and S with straight line on ternary phase diagram.
- Connect S and R_N with straight line on ternary phase diagram.
- Search for E_1 point on equilibrium curve such that:
  - E_1R_N line intersects with SF line at mixing point with the maximum possible distance from F (minimum possible distance from S. This mixing point is denoted as M_{max}.
  - E_1F line must intersects with SR_N at the operating point denoted as P_{max}.

→ If E_1 is given, similar search approach to locate R_N can be followed.

Remark. With S_{max}/F, the required number of theoretical stages:

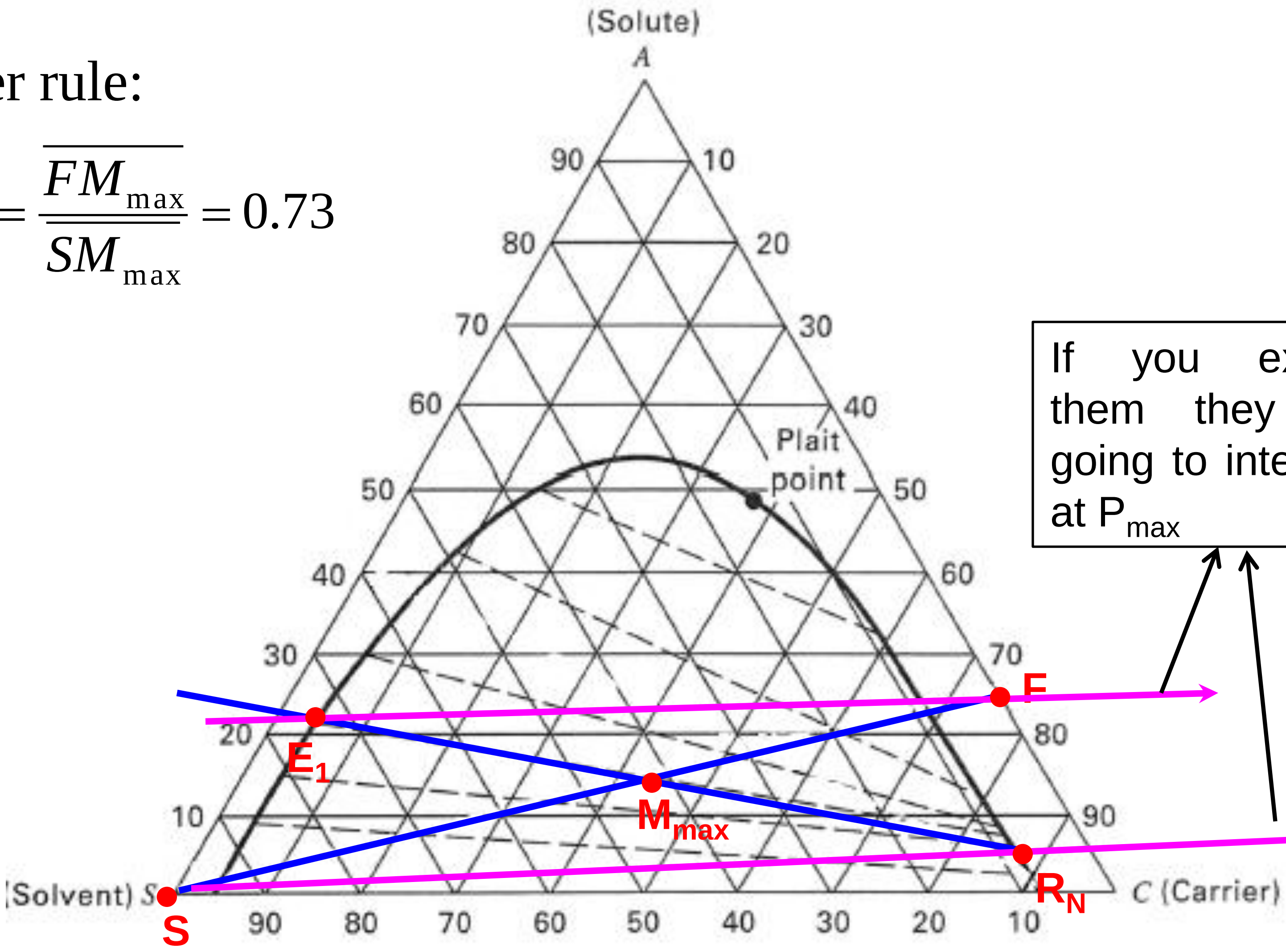
$N \rightarrow 1$

Counter-current graphical solution

Example. Determine S_{max}/F for the system of previous example

- Lever rule:

$$\frac{S_{max}}{F} = \frac{\overline{FM_{max}}}{\overline{SM_{max}}} = 0.73$$



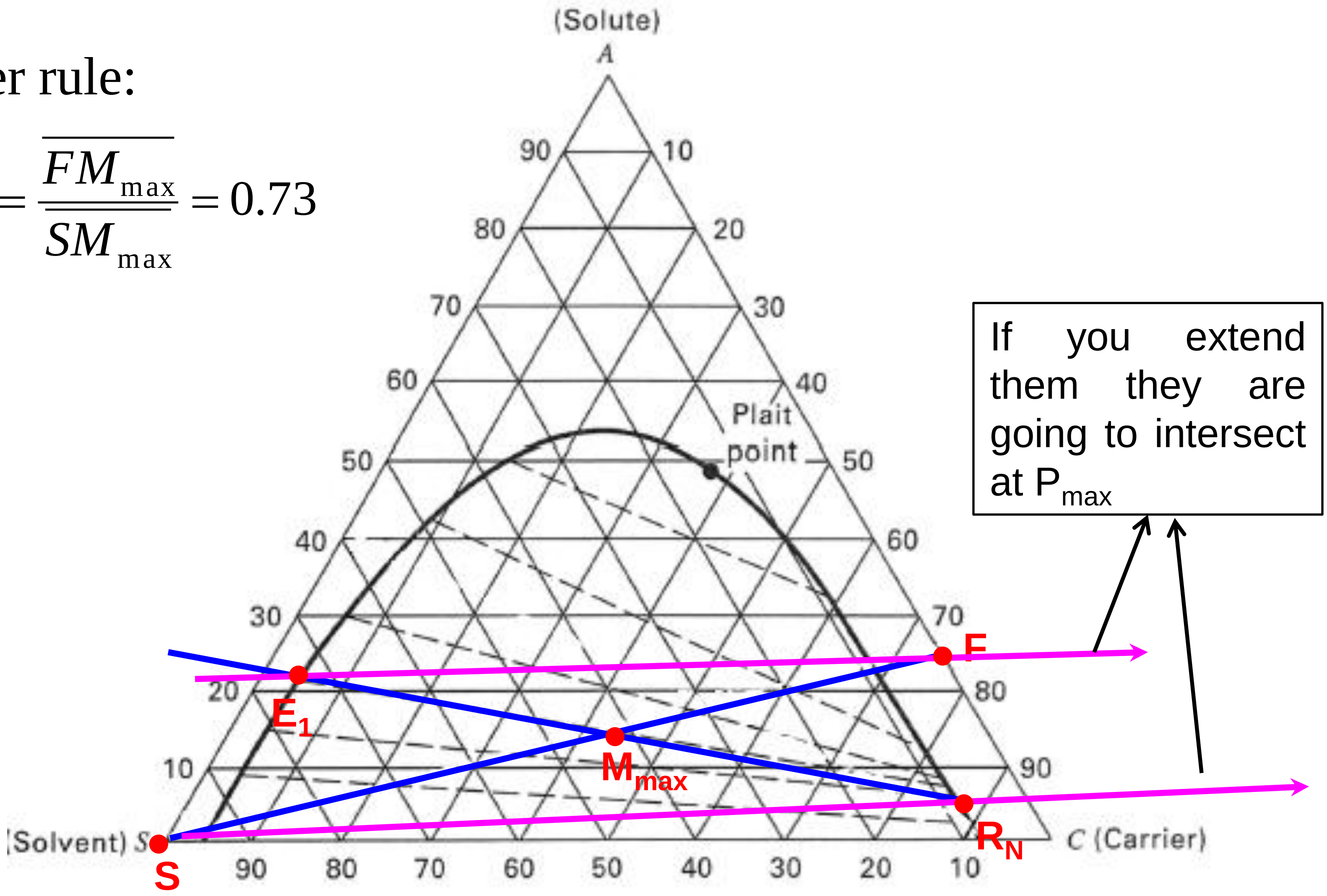
If you extend them they are going to intersect at P_{max}

Counter-current graphical solution

Example. Determine S_{max}/F for the system of previous example

- Lever rule:

$$\frac{S_{\max}}{F} = \frac{\overline{FM_{\max}}}{\overline{SM_{\max}}} = 0.73$$



If you extend them they are going to intersect at P_{max}

Counter-current graphical solution

Remark.

- For extraction operation: $\left(\frac{S_{\min}}{F}\right) < \left(\frac{S}{F}\right) < \left(\frac{S_{\max}}{F}\right)$ or $S_{\min} < S < S_{\max}$
- A reasonable value : $S \approx 1.5 S_{\min}$

Example. For previous example, check if the flow rate of solvent (S=28 kg/h) is reasonable for the given feed flow rate (F=112 kg/h).

$$S_{\min}/F = 0.12 \rightarrow S_{\min} = 0.12 F = (0.12)(112) = 13.4\,\mathrm{kg/h}$$

A reasonable value: $\mathrm{S} \approx 1.5 S_{\min} = 20.1\,\mathrm{kg/h}$

Thus, the solvent flow rate of (S=28 kg/h) is a little bit higher than the reasonable one.

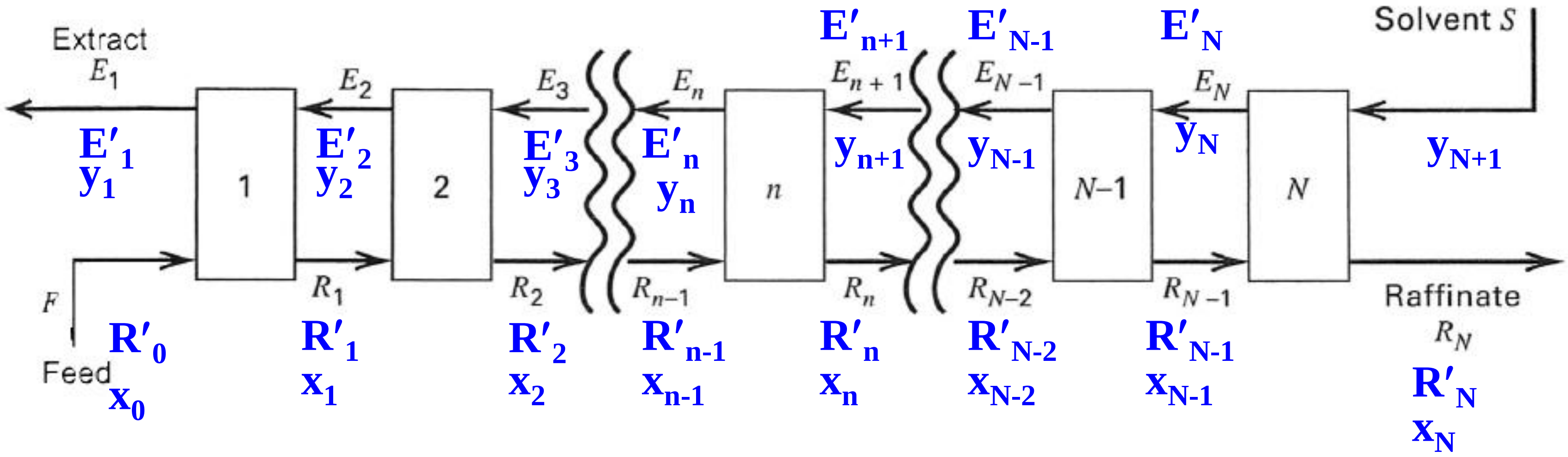
Exercise. Resolve the previous exampleS= 20 kg/h to calculate the required number of theoretical stages.

Counter-current stage extraction with immiscible liquids

- The carries (B) of solute (A) is immiscible in solvent (S).
- In other words, solute is the only component distributed in extract and raffinate.
- This means that the raffinate has a binary mixture of A and B and the extract has binary mixture of A and S.
- In such ternary coordinates is not helpful.
- As in distillation and absorption, the equilibrium xy diagram is used instead.
- Let

  x :solute concentration in raffinate
  y: solute concentration in extract
  E′: flow rate of solvent in the extract streams
  R′: flow rate of carrier B in the raffinate streams

Counter-current stage extraction with immiscible liquids



- **Applying mass balance for carries B on each stage gives:**

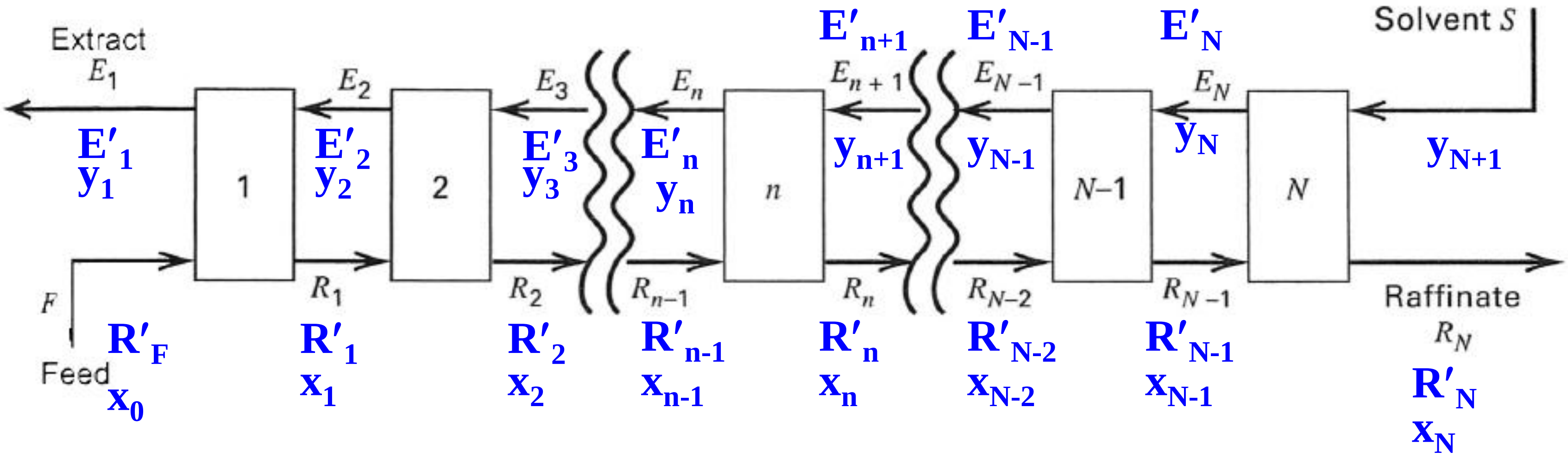
$$R'_0 = (1 - x_0)F = R'_1 = R'_2 = ... = R'_2 = \text{constant} = R'$$

- **Applying mass balance for solvent S on each stage gives:**

$$E'_1 = E'_1 = E'_2 = ... = E'_N = (1 - y_{N+1})S = \text{constant} = E'$$

If pure solvent is used: $y_{N+1} = 0 \Rightarrow \mathrm{E}' = \mathrm{S}$

Counter-current stage extraction with immiscible liquids



- **Applying mass balance for solute A over stages: 1→n :**

$$R'\left(\frac{x_0}{1-x_0}\right)+E'\left(\frac{y_{n+1}}{1-y_{n+1}}\right)=R'\left(\frac{x_n}{1-x_n}\right)+E'\left(\frac{y_1}{1-y_1}\right)$$

Counter-current stage extraction with immiscible liquids

- Let us now define:
  - The **solute-to-carrier mass ratio** in the raffinate:

$$X = \frac{x}{1-x} [=] \frac{\text{kg solute A}}{\text{kg carrier B}}$$

  - and **solute-to-solvent mass ratio** in the extract:

$$Y = \frac{y}{1-y} [=] \frac{\text{kg solute A}}{\text{kg solvent S}}$$

- **Using such mass ratio, the overall mass balance for solute A over stages: 1→n becomes:**

$$E'Y_{n+1} = R'X_n + E'Y_1 - R'X_0$$

- **Dividing by E′ gives:** $\boxed{Y_{n+1} = \frac{R'}{E'} X_n + Y_1 - \frac{R'}{E'} X_0}$

Operating line equation

Counter-current stage extraction with Immiscible liquids

$$\boxed{Y_{n+1} = \frac{R'}{E'} X_n + Y_1 - \frac{R'}{E'} X_0}$$

- **For the above operating line equation:**
  - Plot of Y_{n+1} versus X_n on X-Y diagram gives a straight line with:

$$\boxed{\text{Slope} = \frac{R'}{E'}} \qquad \boxed{\text{Intercept} = Y_1 - \frac{R'}{E'} X_0}$$

  - For certain feed flow rate, the minimum solvent flow rate, S_{min} must correspond to the maximum allowable slope:

$$E'_{\min} = \frac{R'}{\text{Slope}_{\max}}$$

$$\text{But } (1 - y_{N+1})S = E' \Rightarrow S_{\min} = \frac{E'_{\min}}{(1 - y_{N+1})} = \frac{R'}{(1 - y_{N+1})(\text{Slope}_{\max})}$$

  - If x and y are quite small(dilute solutions): $X \approx x \; ; Y \approx y$

Counter-current stage extraction with Immiscible liquids

Example. An inlet water solution of 100 kg/h containing 0.01 wt fraction nicotine (A) in water is stripped with kerosene stream of 200 kg/h containing 0.0005 wt fraction nicotine in a countercurrent stage tower. The water and kerosene are essentially immiscible in each other. It is desired to reduce the concentration of exit water to 0.0010 wt fraction nicotine. Determine.

a. Number of the theoretical stages needed.

b. The minimum solvent rate.

The equilibrium data are as follows:

x	y
0.001010	0.000806
0.002460	0.001959
0.005000	0.004540
0.007460	0.006820
0.009880	0.009040
0.020200	0.018500

Counter-current stage extraction with Immiscible liquids

Solution.

- Solute (A) ≡ Nicotine; Carrier (B) ≡ Water
  Solvent (S) ≡ Kerosene

$$\boxed{\begin{array}{l} F = 100 \text{ kg/h} ; S = 200 \text{ kg/h} \\ x_0 = 0.01 ; x_N = 0.001 ; y_{N+1} = 0.0005 \end{array}}$$

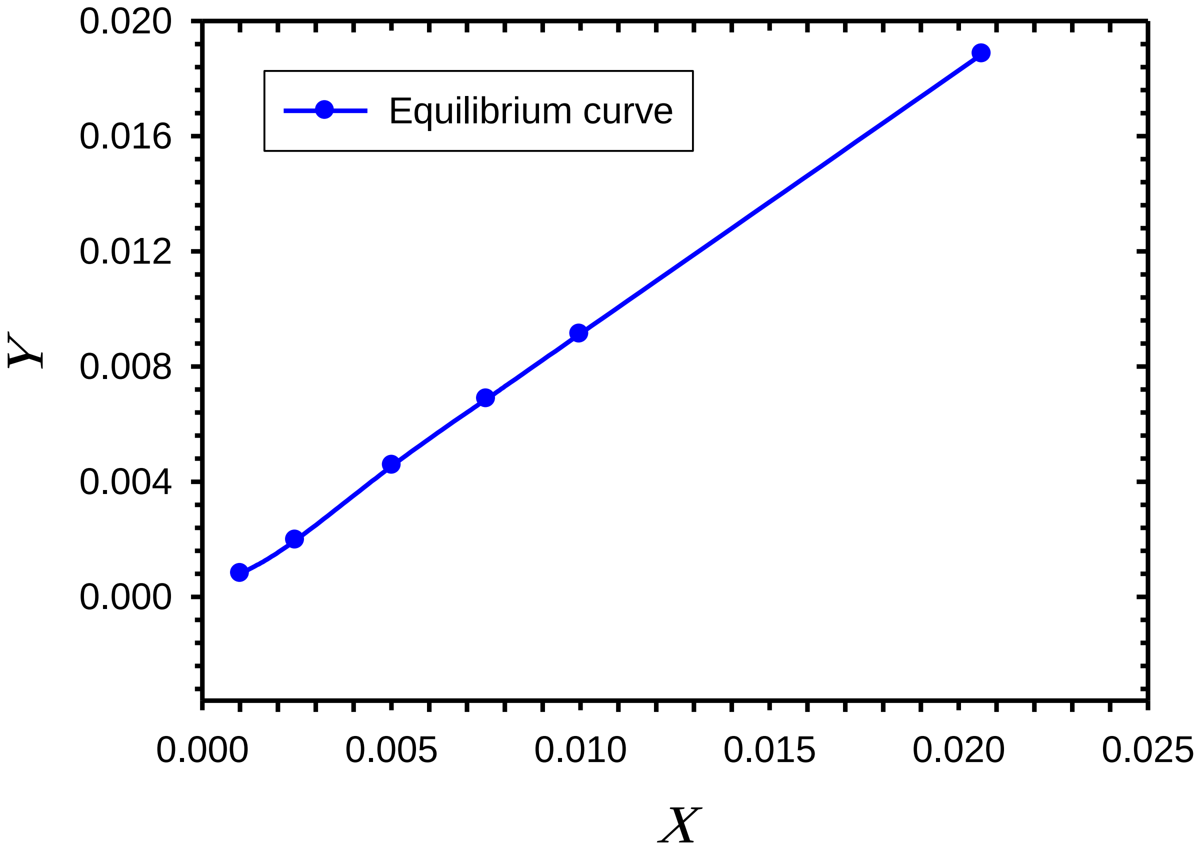
- Re-describe the equilibrium data in terms of solute-to-carrier mass ratio (X) and solute-to-solvent mass ratio in the raffinate

$X = x/(1-x)$	$Y = y/(1-y)$
0.001011	0.000807
0.002466	0.001963
0.005025	0.004561
0.007516	0.006867
0.009979	0.009122
0.020616	0.018849

- Draw XY equilibrium curve

Counter-current stage extraction with Immiscible liquids

Solution.



Counter-current stage extraction with Immiscible liquids

Solution.

$$\boxed{\begin{array}{l} F = 100 \text{ kg/h} \,;\, S = 200 \text{ kg/h} \\ x_0 = 0.01 \,;\, x_N = 0.001 \,;\, y_{N+1} = 0.0005 \end{array}}$$

- **Draw the operating line:**

1. Calculate flow rate of water: $R' = (1 - x_0)F = 99.0 \text{ kg/h}$
2. Calculate flow rate of solvent (Kerosene):

$$E' = (1 - y_{N+1})S = 200\,(1 - 0.0005) = 199.9 \text{ kg/h}$$

3. Find the slope of operating line equation:

$$\text{Slope} = R'/E' = 0.50$$

4. Calculate:

$$X_0 = x_0/(1 - x_0) = 0.01$$

$$X_N = x_N/(1 - x_N) = 0.001$$

$$Y_{N+1} = y_{N+1}/(1 - y_{N+1}) = 0.0005$$

Counter-current stage extraction with Immiscible liquids

Solution.

5. Determine Y_1. The point $(X_N,Y_{N+1})=(0.001,0.0005)$ must lies on the operating line, thus

$$Y_{N+1}=\frac{R'}{E'}X_N+Y_1-\frac{R'}{E'}X_0$$

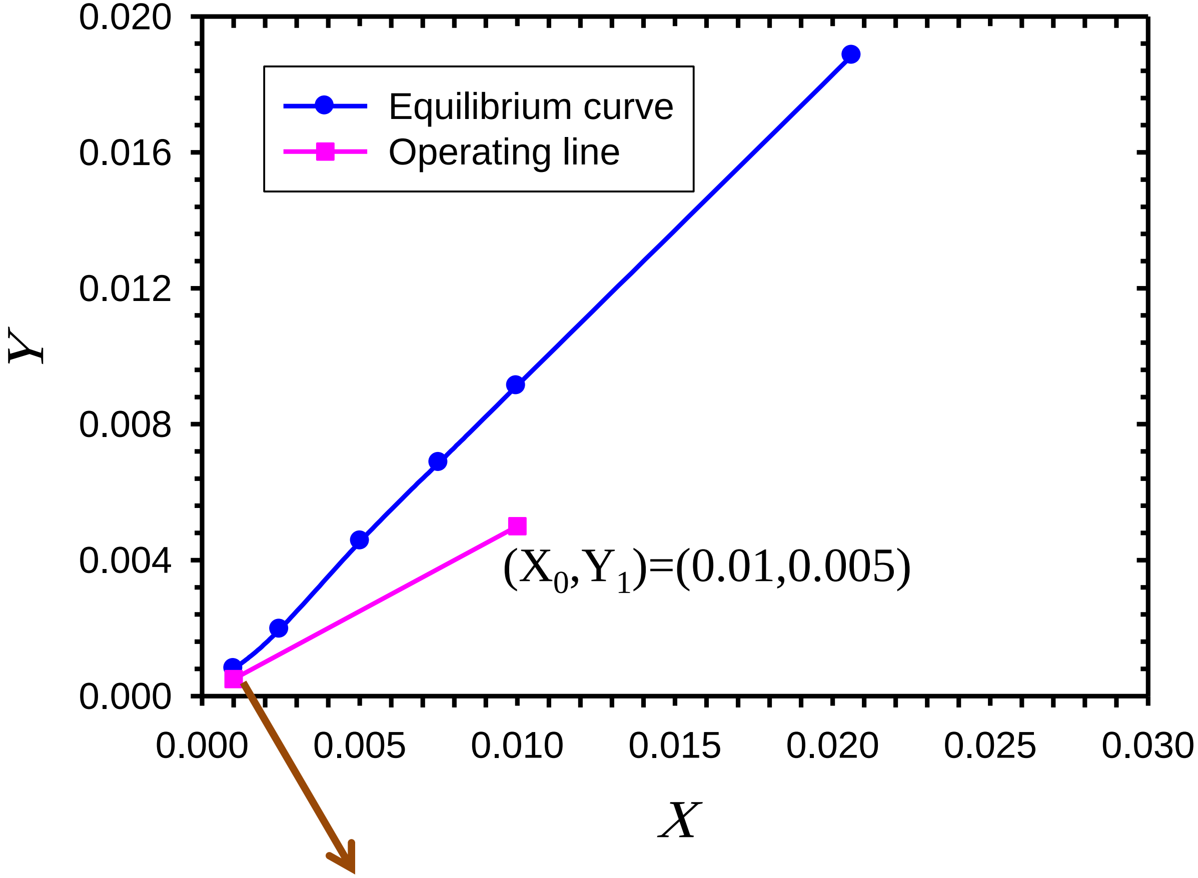
$$0.0005=(0.50)(0.001)+Y_1-(0.50)(0.01)\rightarrow Y_1=0.005$$

The point $(X_0,Y_1)=(0.01,0.005)$ must also must lies on the operating line.

6. With the two point: $(X_N,Y_{N+1})=(0.001,0.0005)$ and $(X_0,Y_1)=(0.01,0.005)$ we can now draw the operating line.

Counter-current stage extraction with Immiscible liquids

Solution.

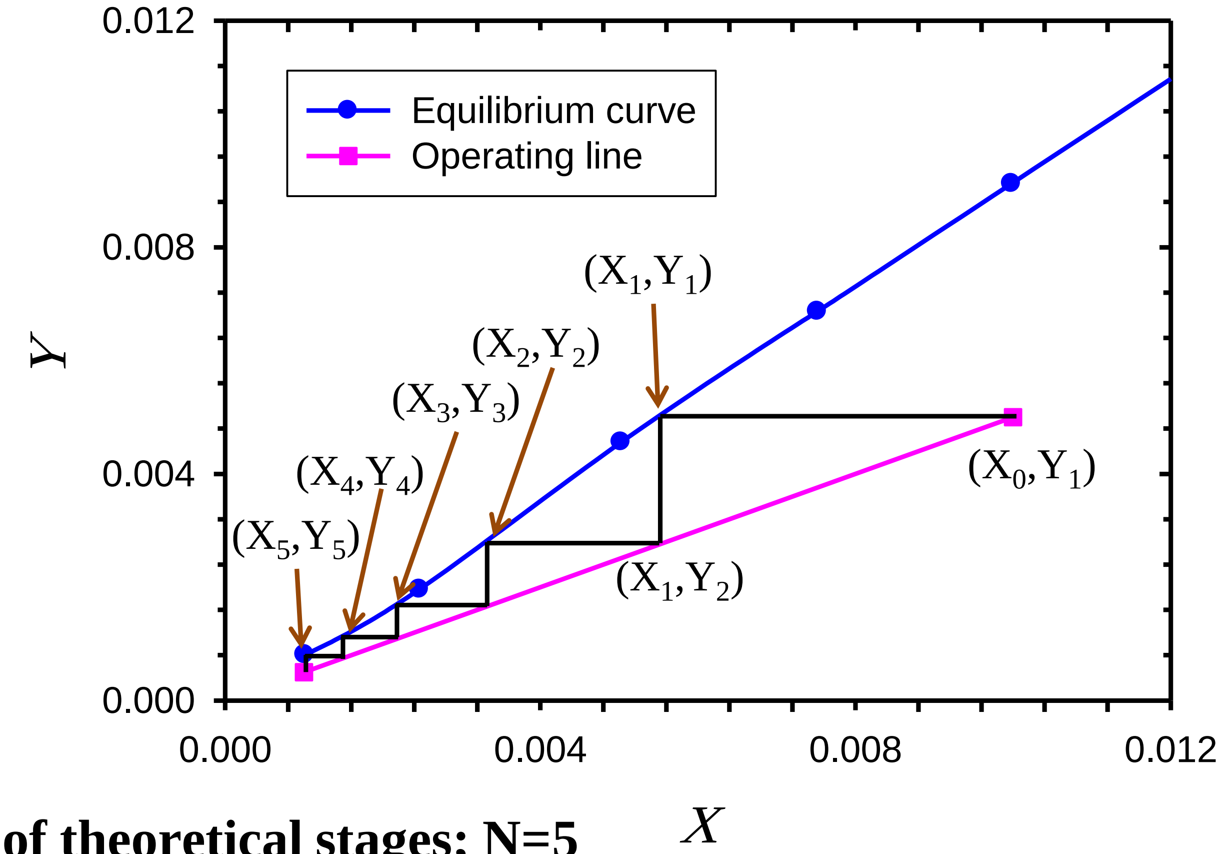


$(X_N, Y_{N+1}) = (0.001, 0.0005)$

Counter-current stage extraction with Immiscible liquids

Solution.

- The number of theoretical stages are stepped off, as in distillation and absorption, from as (X_0,Y_1) to (X_N,Y_{N+1}):

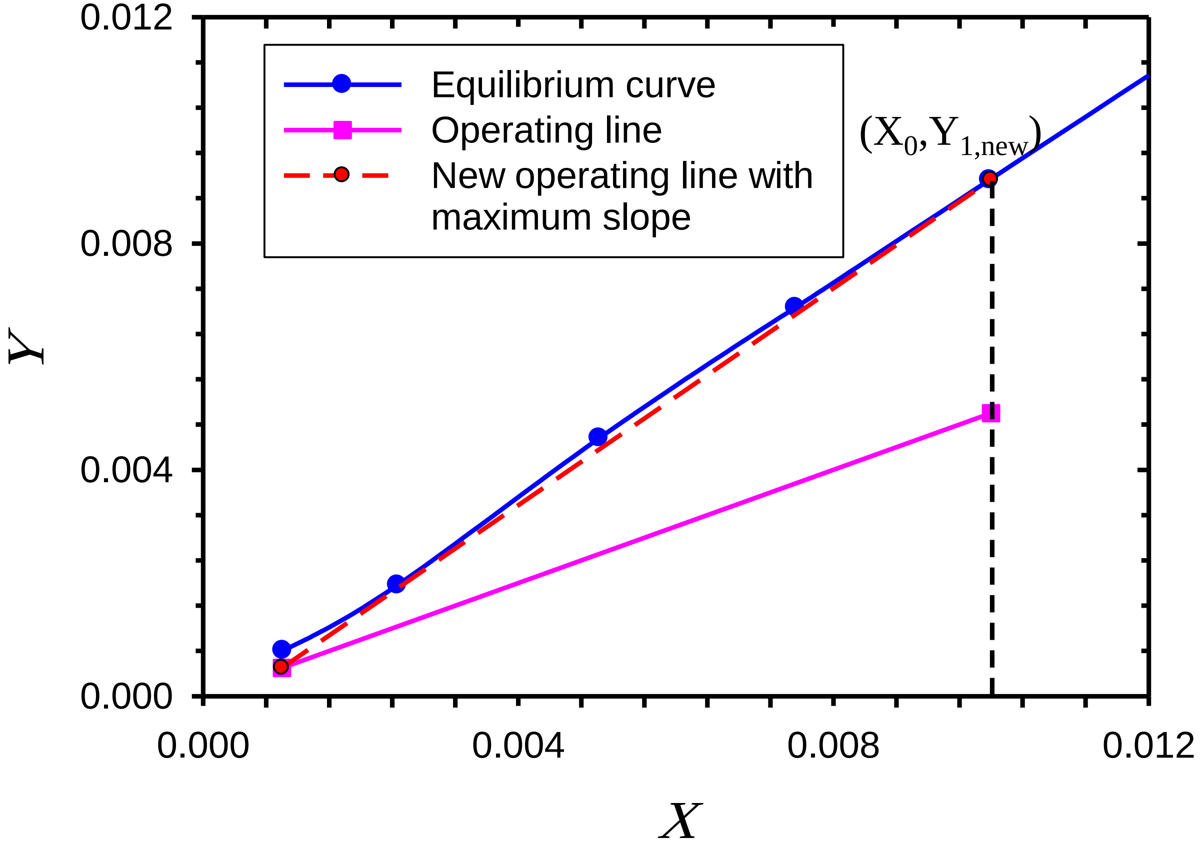


a) The number of theoretical stages; N=5

Counter-current stage extraction with Immiscible liquids

Solution.

b) Minimum solvent flow → Maximum allowable slope



$$Slope_{max} = \frac{Y_{1,new} - Y_{N+1}}{X_0 - X_N} = \frac{0.009122 - 0.0005}{0.01 - 0.001} = 0.96$$

Counter-current stage extraction with Immiscible liquids

Solution.

b) Minimum solvent flow → Maximum allowable slope

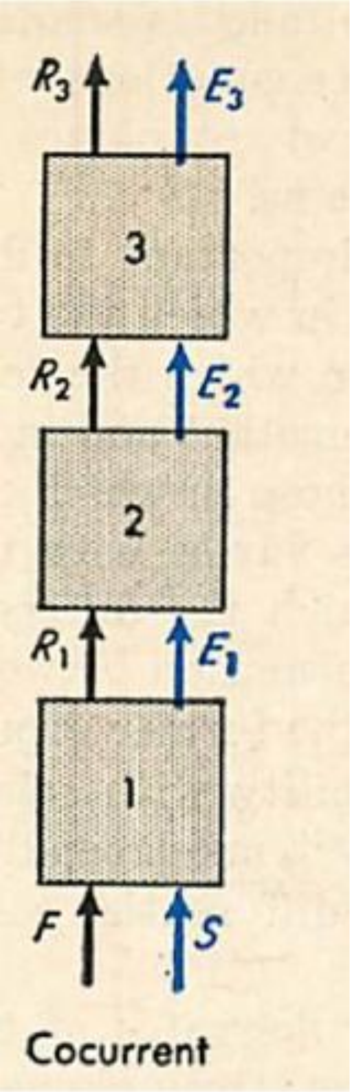
$$\text{Slope}_{\max} = 0.96 = \frac{R'}{E'_{\min}} = \frac{99.0}{E'_{\min}} \rightarrow E'_{\min} = 103.1 \text{ kg/h}$$

$$E'_{\min} = 103.1 = \text{S}_{\min}(1 - y_{\text{N-1}}) = \text{S}_{\min}(1 - 0.0005)$$

$$\rightarrow \text{S}_{\min} = 103.1 \text{kg/h}$$

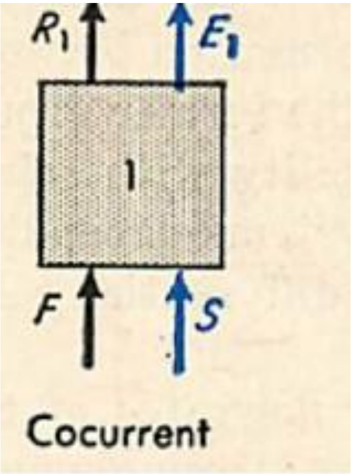
- If reasonable solvent flow rate; $S=1.5S_{\text{min}}$, for such process we should use around 155 kg/h. Try to find number of stages with this solvent flow rate.

Cocurrent one stage extraction



$$\boxed{M = S + F = R_1 + E_1 = R_2 + E_2 = \ldots\ldots.. = R_N + E_N}$$

Obviously, always there will be only one theoretical stage operating in cocurrent mode.



Converting theoretical stages to actual equipment size

Assume the counter-current stage extraction needs 6 theoretical stages to achieve some required recovery:

- This does not mean we require 6 mixer-settlers (though we could do that, but costly).
- It means we need an extraction column which has equivalent operation of 6 counter-current mixer-settlers that fully reach equilibrium.
- At this point, we resort to correlations and vendor assistance.
- Vendors: provide **HETS** = Height Equivalent to a Theoretical Stage.
- Use that to size the extraction column:

$$\boxed{H = \frac{\text{HETS} \times \text{N}}{\text{Stage efficiency}}}$$

Where H is the height of extraction tower

Converting theoretical stages to actual equipment size

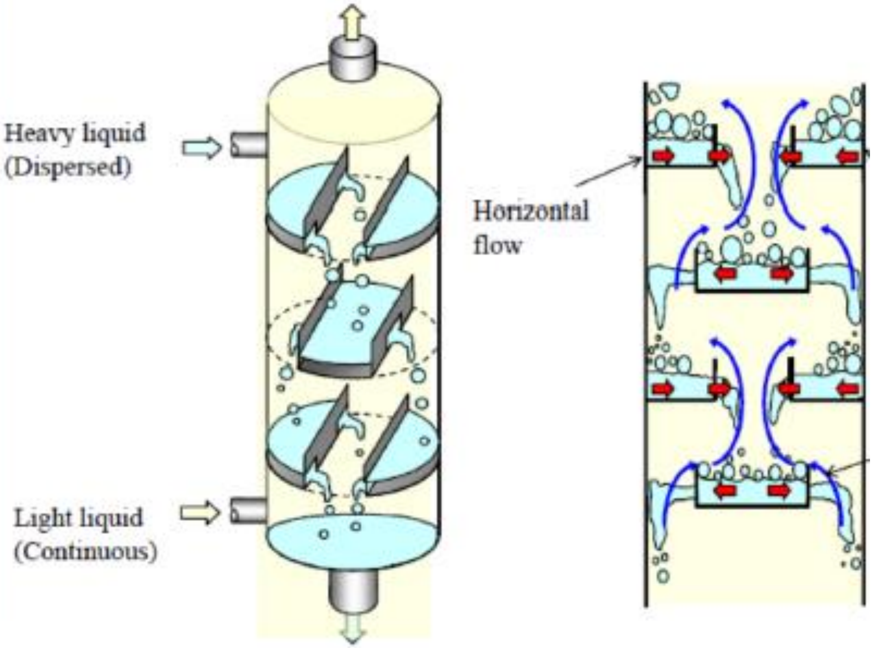


Figure 2: Concept and Flow of WINTRAY System

[WINTRAY (Japanese company; newly patented design)]

Converting theoretical stages to actual equipment size

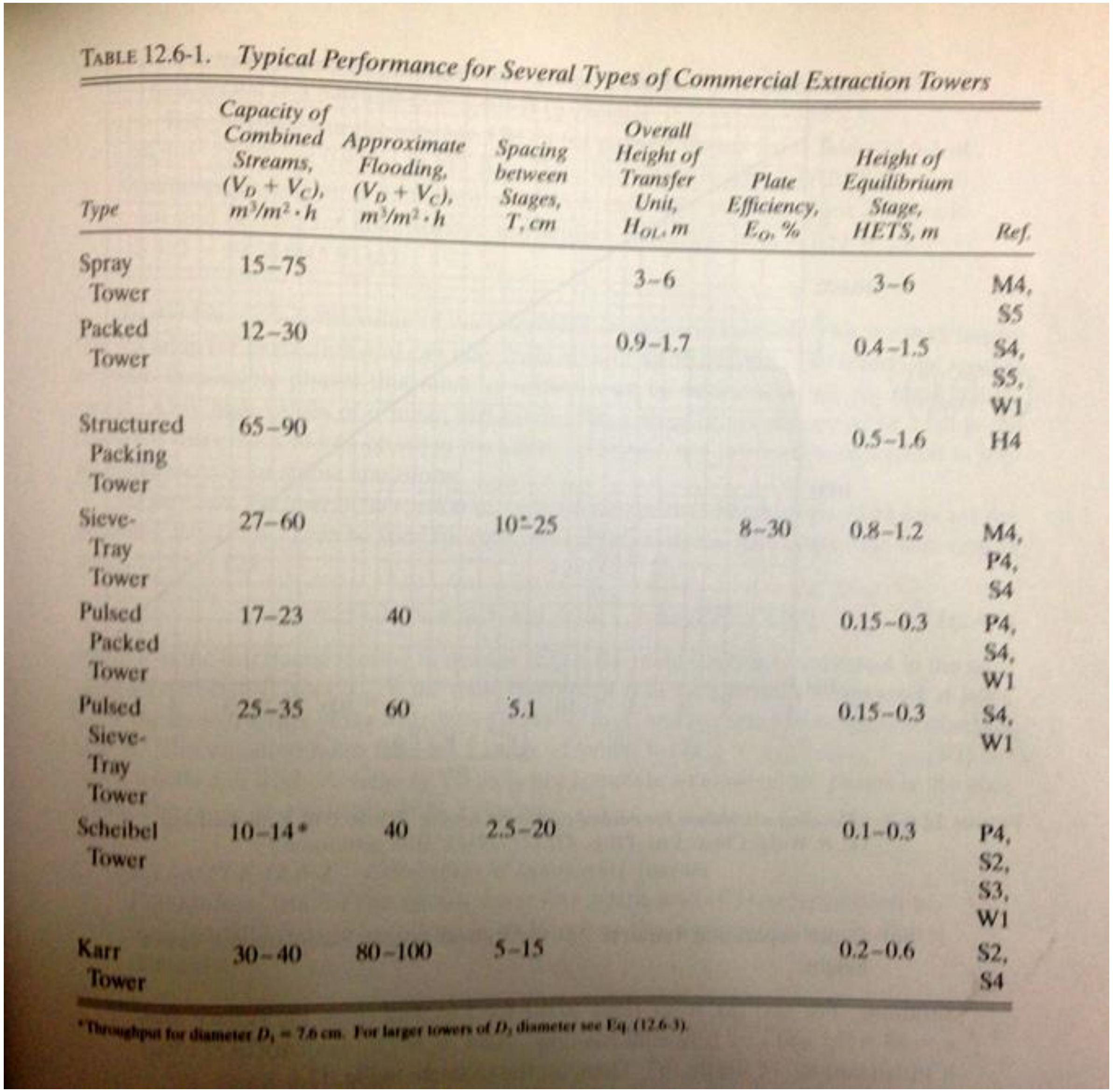


TABLE 12.6-1. *Typical Performance for Several Types of Commercial Extraction Towers*

Type	Capacity of Combined Streams, $(V_D + V_C)$, $m^3/m^2 \cdot h$	Approximate Flooding, $(V_D + V_C)$, $m^3/m^2 \cdot h$	Spacing between Stages, T, cm	Overall Height of Transfer Unit, H_{OL}, m	Plate Efficiency, E_O, %	Height of Equilibrium Stage, HETS, m	Ref.
Spray Tower	15–75			3–6		3–6	M4, S5
Packed Tower	12–30			0.9–1.7		0.4–1.5	S4, S5, W1
Structured Packing Tower	65–90					0.5–1.6	H4
Sieve-Tray Tower	27–60		10–25		8–30	0.8–1.2	M4, P4, S4
Pulsed Packed Tower	17–23	40				0.15–0.3	P4, S4, W1
Pulsed Sieve-Tray Tower	25–35	60	5.1			0.15–0.3	S4, W1
Scheibel Tower	10–14*	40	2.5–20			0.1–0.3	P4, S2, S3, W1
Karr Tower	30–40	80–100	5–15			0.2–0.6	S2, S4

*Throughput for diameter $D_1 = 7.6$ cm. For larger towers of D_2 diameter see Eq. (12.6-3).