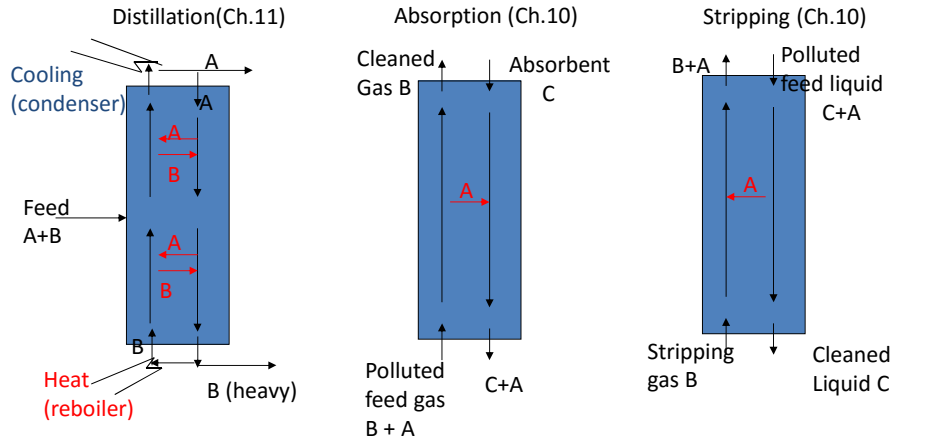
Absorption/Stripping

Gas-liquid separation processes

(Ch. 10)

COUNTER-CURRENT MULTISTAGE CONTACT OF GAS AND LIQUID



- **Two** key components (A+B)
- One feed (normally)
- Two pure products
- Two sections
- Heating and cooling
- Boiling mixture

- **Three** components
  - Inert gas B (e.g., air)
  - Inert liquid C (e.g., water)
  - Solute A (e.g., CO2) goes from gas -> liquid
- Two feeds
- One pure product (at most)
- One section
- Non-boiling mixture
  - so p and T can be set independently; see Exercise

Stripping: Reverse of absorption. Solute A goes from liquid -> gas

Sigurd Skogestad and Jan-Fredrik Arnulf (2013)

«VANNVASK»

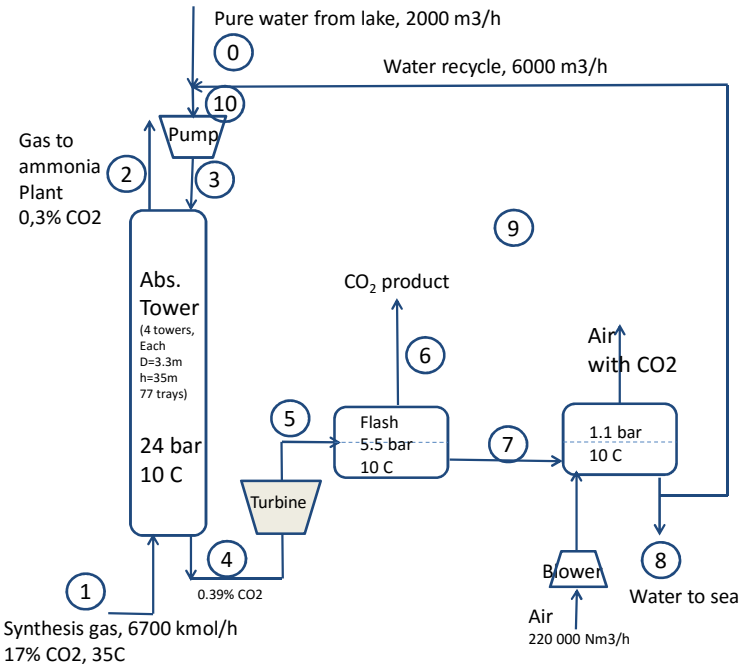
4 identical absorption columns using **pure water** to absorb CO_2 from N_2/H_2 in ammonia plant.

The only one in the world, as far as I know

Originally, the CO_2 was used to make urea but now the CO_2 is the main supply of «brus» in Norway and Sweden.

Designed about 1963 by Ingulf Skogestad, Norsk Hydro Engineering Office, Oslo.
Built at Herøya (Porsgrunn) about 1967 (Norsk Hydro ammonia plant NII, now Yara)

EXAM, Dec. 2016



Revidert SIS. 24. jan. 2017

Remove water from natural gas

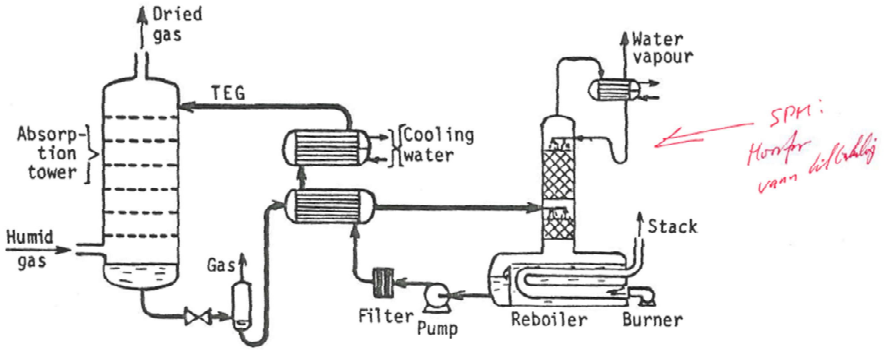
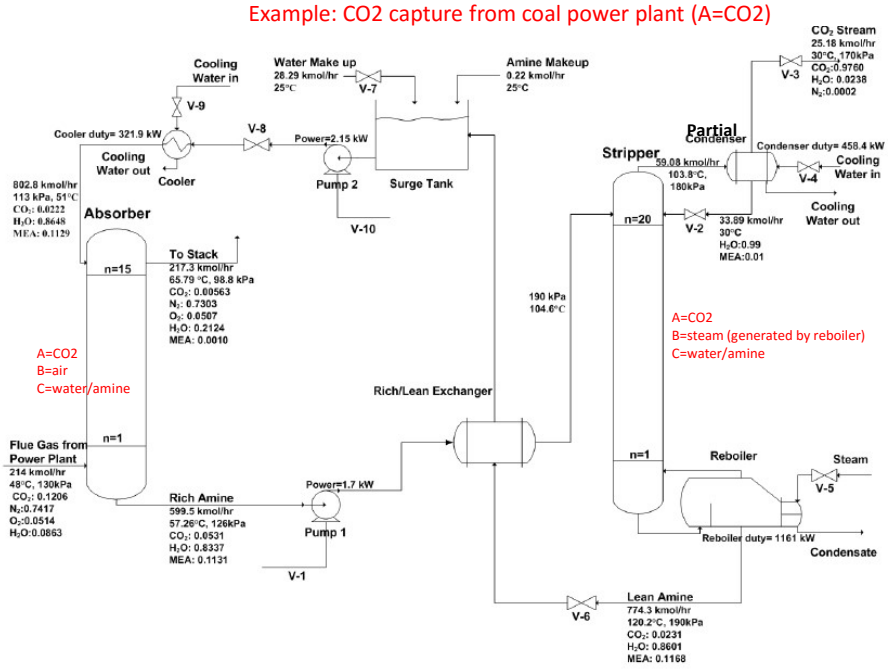


Figure 2.61 Triethylene glycol (TEG) regenerator

Example: CO2 capture from coal power plant (A=CO2)



Absorption/stripping

- Two solution approaches

1. Equilibrium-stage model (Ch. 10.3). Find N
   - Similar to distillation
   - (A) Graphical: McCabe-Thiele (any VLE)
   - (B) Analytical: Kremser (VLE: Assume dilute solution. 1) Henry's law, y = mx, +2) assume L/V constant)
2. Mass transfer model, (Ch. 10.4+10.6).
   - Find A=aSz (A=interfacial area)
   - Similar to heat transfer in heat exchanger
     - (A) Graphical: Must combine with integration
     - (B) Analytical (Henry's law): Log-mean driving, $(y-y^*)_{LM}$

Graphical method (McCabe-Thiele) for absorption/stripping

- Equilibrium line usually straight line because of dilute solution
- Operating line goes through end points
  - These point are <u>not</u> on the diagonal*
  - Usually straight operating line because of dilute mixture

- *Note: For distillation the operating lines go through (x_D,x_D) and (x_B,x_B), which are on the diagonal.
  - Reason: Reflux/boilup generates "feed stream" with same composition as product.

Flows [mol/s]

Flows generally vary through the column

- L = L'/(1-x)
- V = V'/(1-y)

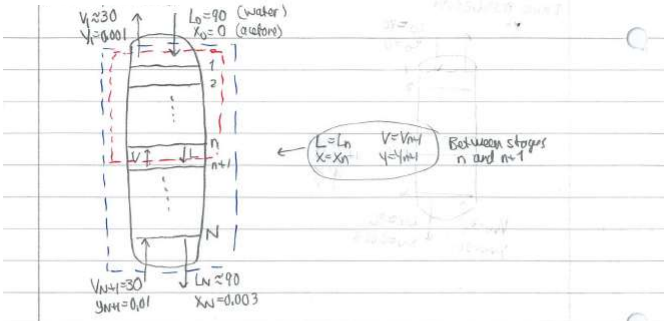
x, y = mole fractions of component being transferred (A)
L' = inert liquid molar flow (C)
V' = inert vapor molar flow (B)

For dilute mixtures we can assume (1-x)=1 and (1-y)=1 so L and V are constant through the column.

Example 10.3-2. Absorption of Acetone in a Countercurrent Stage Tower
It is desired to absorb 90% of the acetone in a gas containing 1.0 mol % acetone in air in a countercurrent stage tower. The total inlet gas flow to the tower is 30.0 kg mol/h, and the total inlet pure water flow to be used to absorb the acetone is 90 kg mol H_2O/h. The process is to operate isothermally at 300 K and a total pressure of 101.3 kPa. The equilibrium relation for the acetone (A) in the gas–liquid is $y_A = 2.53x_A$. Determine the number of theoretical stages required for this separation.

m=2.53, L/V=3
L=90 kmol/h, V=30 kmol/h

From last time:



1. Overall material balance for A (around ends)

In A = Out A

$y_{N+1}V_{N+1} + x_0L_0 = y_1V_1 + x_NL_N$ → x_N=0.003

2. Balance for A between stages (here: over top, but could do btm instead)

In A = Out A

$yV + x_0L_0 = y_1V_1 + xL$

⇒ $y = \frac{L}{V}x + \frac{y_1V_1 - x_0L_0}{V}$ ← operating line. Gives y=f(x) between stages

Note: 1) Operating line goes through points $(x,y)=(x_0,y_1)$ (TOP) and $(x,y)=(x_N,y_{N+1})$ (BTM)!!

2) Operating line is straight if L and V are constant.

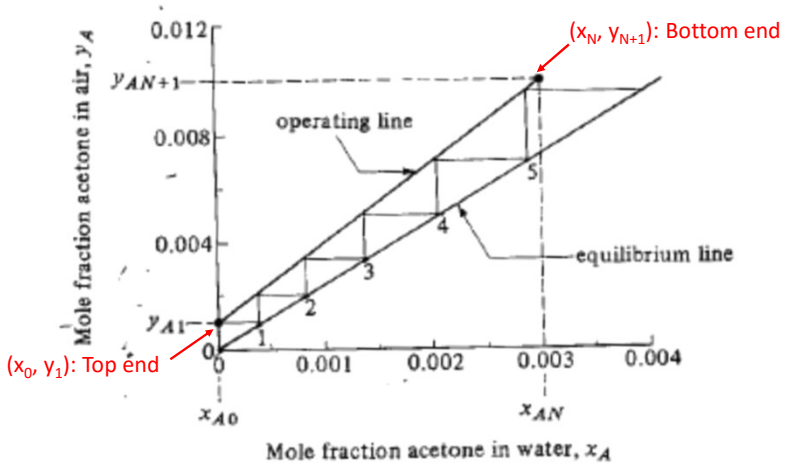


FIGURE 10.3-4. *Theoretical stages for countercurrent absorption in Example 10.3-2.*

m=2.53, L/V=3
L=90 kmol/h, V=30 kmol/h
Solution: Need N=5.2 (approx.)

Minimum flows
Absorption: L_{min}
Stripping: V_{min}

- Absorption: Pinch (equilibrium) in bottom where vapor feed enters
- Stripping: Pinch (equilibrium) in top where liquid feed enters
- Corresponds to N=∞

L_{min}: Back to example

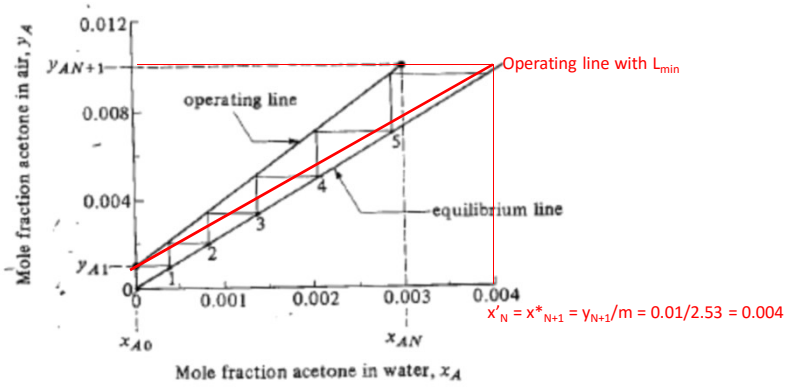


FIGURE 10.3-4. *Theoretical stages for countercurrent absorption in Example 10.3-2.*

From figure: $L_{min}/V = (y_{N+1} - y_1)/(x^*_{N+1}-x_0) = (0.01-0.001)/(0.004-0) = 2.277$
So: $L_{min} = 30*2.277 = 68.3$ kmol/h

* = in equlibrium with the other phase
Example: x^*_{N+1} is (imaginary) liquid composition in equlibrium with y^*_{N+1}

L_{min}: «Alternative» derivation

- From overall mass balance (assuming constant flows)

  $V\, y_{N+1} + L\, x_0 = V\, y_1 + L\, x_N$

- Assume equlibrium in bottom (pinch)

  $x_N = x^*_{N+1} = y_{N+1}/m = 0.0040$

- Get min. reflux. $L_{min}/V = (y_{N+1} - y_1)/(x^*_{N+1}-x_0)$

Minimum stages, N_{min}

- No such thing for absorption/stripping (N_{min}=0)
- Example: 1 stage is OK if we increase flows enough
  - Absorption: increase L enough
  - Stripping: increase V enough

N=1: Back to example

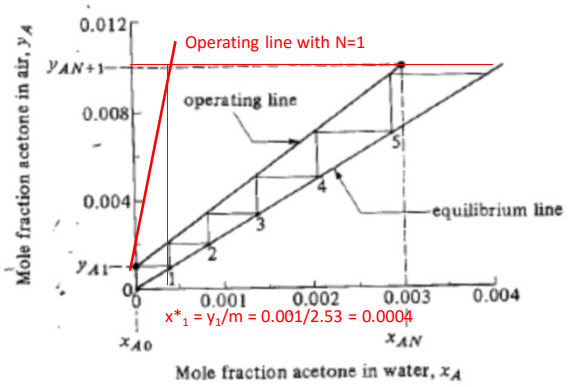
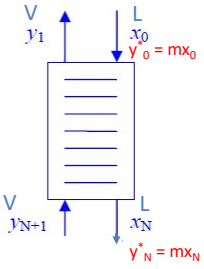


FIGURE 10.3-4. *Theoretical stages for countercurrent absorption in Example 10.3-2.*

From figure: $L/V = (y_{N+1} - y_1)/(x^*_1 - x_0) = (0.01-0.001)/(0.0004-0) = 22.77$
So: $L = 30*22.77 = 683$ kmol/h

Kremser equations

- Analytical solution for case with straight equilibrium line (y=mx) and L/V constant (dilute mixtures)
- A = (L/V)/m – absorption factor = ratio of slopes for operating and equilibrium lines
  - Absorption: want A large
  - Stripping: Want A small



Book eq. (10.3-25):

$$\frac{y_{N+1}-y_1}{y_{N+1}-y_0^*} = \frac{A^{N+1}-A}{A^{N+1}-1}$$

Alternative simpler form (Sigurd):

$$\frac{y_{N+1}-y_N^*}{y_1-y_0^*} = A^N$$

Get:

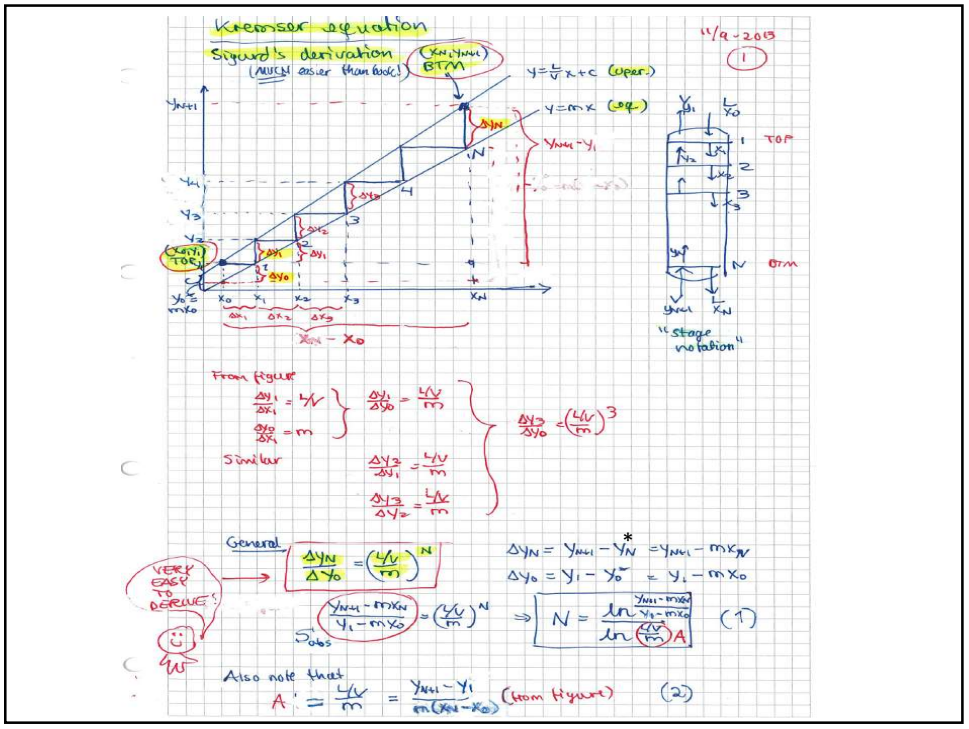
$$N = ln\frac{y_{N+1}-y_N^*}{y_1-y_0^*}/lnA$$

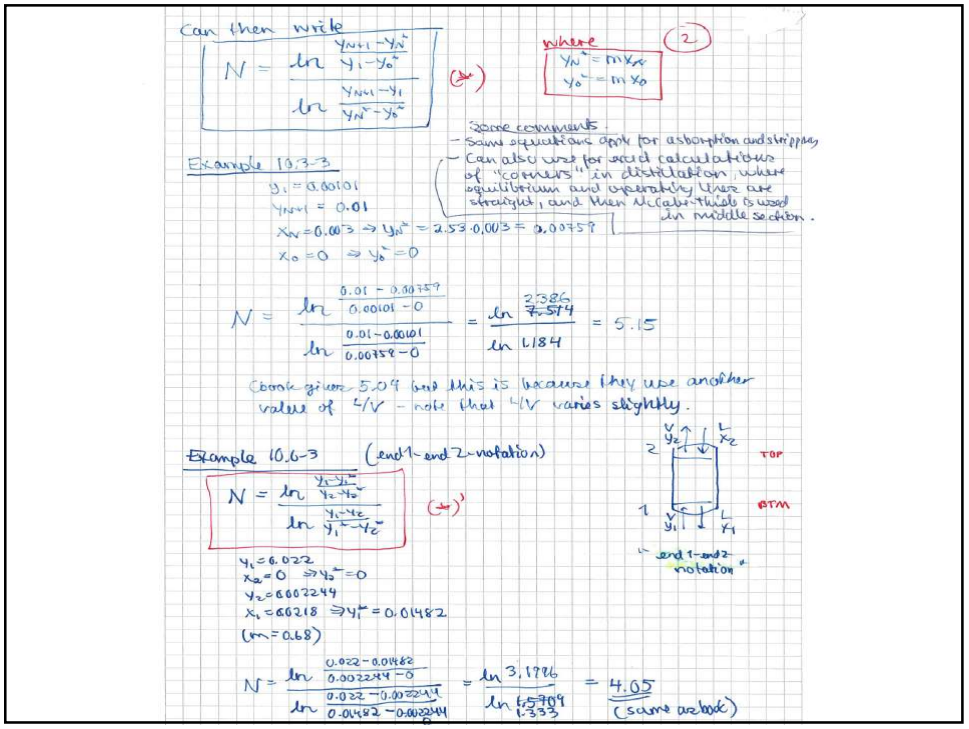
Note. Can write:

$$A = \frac{y_{N+1}-y_1}{y_N^*-y_0^*}$$

Proof: Book or much simpler graphical proof from Sigurd (next page)

- = in equilibrium with other phase (could be imaginary composition, like y^*_0)


Kremsor equation
Stigurd's derivation (MUCH easier than book!)
(xN, yN+1) BTM
(x1, y1) TOP
y = L/V x + c (oper.)
y = mx (eq.)
yN+1 - y1
ΔyN
ΔyN-1
Δy3
Δy2
Δy1
Δy0
xN - x0
TOP
BTM
"stage notation"
From figure
Similar
General
VERY EASY TO DERIVE!
Also note that
A = L/mV
(from figure)
(1)
(2)


Can then write
(1)
where
(2)
Some comments
- Same equations apply for absorption and stripping
- Can also use for word calculations of "counters" in distillation where equilibrium and operating lines are straight, and then N Calculations is used in middle section.
Example 10.3-3
y1 = 0.00101
yN+1 = 0.01
xN = 0.003 ⇒ yN* = 2.53 · 0.003 = 0.00759
x0 = 0 ⇒ y0* = 0
N = ln((0.01 - 0.00759)/(0.00101 - 0)) / ln((0.01 - 0.00101)/(0.00759 - 0)) = ln(2.386)/ln(1.184) = 5.15
Book gives 5.09 but this is because they use another value of L/V - note that L/V varies slightly.
Example 10.6-3 (end 1 - end 2 - notation)
(1)'
y1 = 0.022
x2 = 0 ⇒ y2* = 0
y2 = 0.002244
x1 = 0.0218 ⇒ y1* = 0.01482
(m = 0.68)
N = ln((0.022 - 0.01482)/(0.002244 - 0)) / ln((0.022 - 0.002244)/(0.01482 - 0.002244)) = ln 3.1986 / ln 1.5799 = 4.05
(same as book)
TOP
BTM
"end 1 - end 2 notation"

Minimum flows

Example 1: A fun absorption problem

- Dear Absorption and distillation experts: Can you solve this?

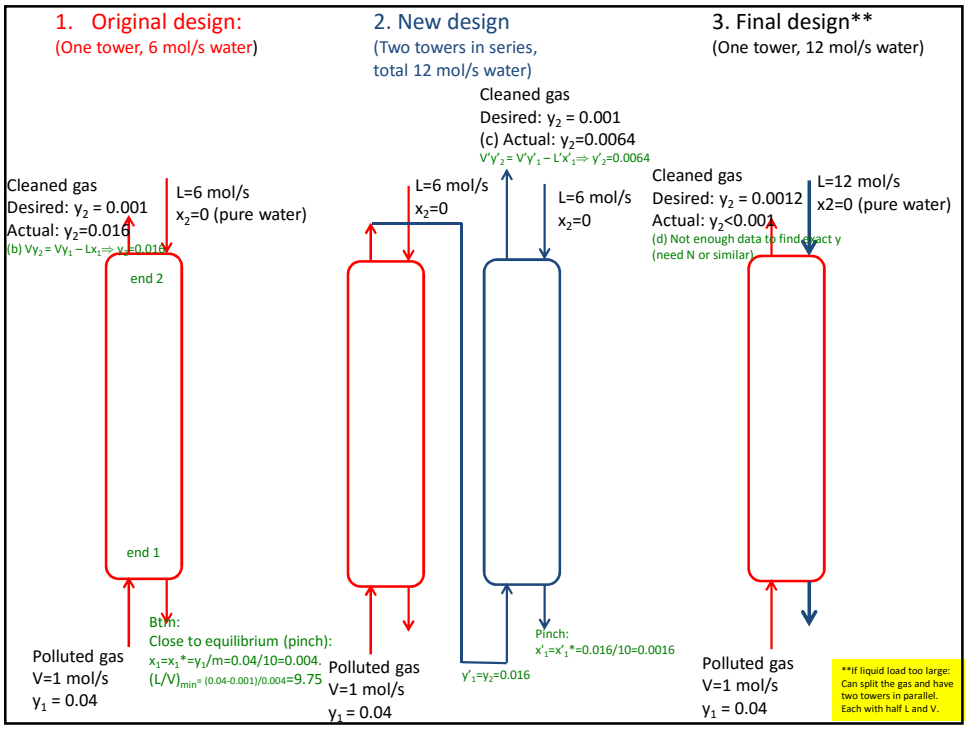
  This problem is based on a true story. My father, Ingulf Skogestad, got his degree as a chemical engineer at NTH in Trondheim in 1948. It was difficult to find good process jobs in Norway at that time, so his career got a boost when he during 1956-61 worked at two ammonia plants that ICI operated in South Africa.

- This is a story from Modderfontein near Johannesburg in 1959 where they had a problem of not reaching the gas spec. in an absorption tower. They build an identical tower no. 2 (in series) but still they could not reach the spec. My father found a solution that worked: He took the absorbent (liquid) from tower 2 and used it in tower 1 (and left tower 2 as a spare). This was at the time explained by the fact that there was too little liquid in tower 1 to get proper wetting of the packing, but I think it may have been a more fundamental mass balance problem.

- To illustrate this I have made the following problem:
  A gas stream of 1 mol/s has 4 mol% of component A, and it should be reduced to 0.1% in an absorption tower with use of 6 mol/s pure water. However, in spite of the tower being very tall, the fraction of A in the gas product is too high (about 1.6% A).
  To reduce the concentration of A, an identical tower 2 is build, where the exit gas from tower 1 (with 1.6% A) is treated with 6 mol/s pure water. However, one still is not able to reach the desired 0.1% A.

  Data: Henry's law for component A: $y = mx$, where $m = H/p = 10$ (p is the column pressure, which is 1 bar).

  (a) Make a flowsheet for the two alternatives.
  (b) Can it be correct that one achieves 1.6% A in tower 1?
  (c) What is the concentration of A in the gas leaving tower 2?
  (d) An engineer always finds a solution! In this case it is to use all the water (12 mol/s) in tower 1 (and leave tower 2 as a spare). Can you explain why this works, and find the exit gas concentration of A in this case.



Minimum flows

Example 2 (Problem 2, exam 2004)

Oppgave 2 STRIPPING (Vekt: 25%)

Avløpsvannet fra en oljeplattform skal renses for aromatiske hydrokarboner ved stripping med luft. Forurensningen tilsvarer 8 mg benzen per liter vann. Målet er å redusere nivået til 0.4 mg benzen per liter vann. Separasjonen skjer ved et totaltrykk på 2 atm.

a) Finn minste luftmengde (m^3 luft per m^3 vann) som trengs til formålet. Hva blir antallet likevektstrinn i dette tilfellet?

Fysikalske data for benzen:

Molvekt: 78

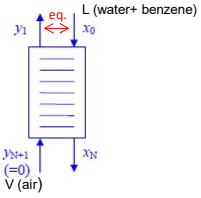
Henrys konstant: $H_i = \frac{y_i}{x_i} p = 383.5$ bar

PROBLEM 2. STRIPPING.
Water from an oil installation contains 8 mg benzene per liter water. It should be reduced to 0.4 mg benzene per liter by stripping with air at 2 atm.
(a) Find the minimum amount of air (m3 air/m3 water). How many equilibrium stages does this require?

SOLUTION. General approach:

(i) Convert everything to molar + (Find product streams from mass balance)

(ii) Minimum flows: This is with infinite no. of stages and it occurs when we have "pinch"
For absorption/stripping: "Pinch" at one end -> product and feed in equilibrium!



Assume: L and V constant through column (dilute-> x and y small)
+ Assume T=300K (not given!) + assume ideal gas
(i) Conversion to molar: "8 mg benzene per liter water".
n_b = 8 mg benzene = 8e-3 g / 78 g/mol = 1.026e-4 mol
n_w = 1 liter water = 1 kg water = 1000g/ 18 g/mol = 55.55 mol
$x_0 = n_b/(n_b+n_x)$ = 1.846 e-6 = 1.846 ppm
Similar: $x_N = x_0$*0.4/8 = 0.05 x_0 = 9.231 e-8 = 0.092 ppm
Infinite number of stages: y_1 and x_0 are in equilibrium -> $y_1 = (H/p)\, x_0$ = (383.5 bar/2bar)*1.846e-6 = 3.54e-4 = 354 ppm
(ii) Can now find minimum air flow from overall material balance:
$Lx_0 + Vy_{N+1} = Lx_N + Vy_1$
Or $L(x_0-x_N) = V(y_1-y_{N+1})$
Or $V = L(x_0-x_N)/(y_1-y_{N+1})$ = L (1.846-0.092)/(354-0) = 0.00495 L [mol/s]
Water: Given L = 1 m3 = 1000 kg = 55.55 kmol
Air: Get V = 0.00495 L = 0.275 kmol = 275 mol = 275 (RT/p) [m3] = 275*8.31*300/2e5 [m3] = 3.43 m3 (gas)

Diffusion (Two-film theory)

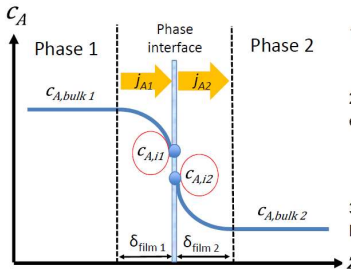
- See lecture no. 3 from Jana
- Mass transfer based on bulk compositiums

  $N_A = K_y (y-y^*) = K_x (x-x^*)$ [mol/s m^2]

  $K_y = m\, K_x$
- Assume dilute mixture, so that drift of film can be neglected:

  $K_y = K'_y$
- To find overall mass transfer coefficient: Add resistances: $1/K_y = 1/k_y + m/k_x$

4) MT across phase interface



ASSUMPTIONS

1. Steady state

$$j_A = j_{A1} = j_{A2}$$

2. Thermodynamic equilibrium at the interface

$$c_{A,i2} = \varphi_A c_{A,i1}$$

Equilibrium coefficient

3. The interphase represents NO resistance for the mass transfer

$$j_A = K_{c1}\left(c_{A,bulk1} - c^{eq}_{A,bulk1}\right)$$

$c^{eq}_{A,bulk\,1}$ - Such a concentration in bulk phase 1, that would be in equilibrium with the current concentration in bulk phase 2. If the current concentration in bulk phase 1 is equal to the equilibrium concentration, there would be NO mas transfer across the interface

K_{c1} - Overall mass transfer coefficient
- Concentration based
- Related to phase 1

$$K_{c1} = \frac{1}{\left(\frac{1}{k_{c1}} + \frac{1}{k_{c2}\,\varphi_A}\right)} \quad \left[\frac{m}{s}\right]$$

Mass transfer approach

- Two-film mass transfer model
- Bulk phases not in equilibrium
- Dilute mixtures (1-x ≈1)
  - Henry's law: $y^*_i = m\,x_i$
  - Constant (molar) flows: L ≈ L' (inert)
- Mass transfer (Non-equilibrium) model
- Get differential equation (dn_A/dz=…) instead of difference equations:
  - Mass transfer: $N_A = K_y(y - y^*)$ [mol A/s, m2]
  - Mass balance over dz: $dn_A = V dy = N_A\,dA$, where $dA = aS dz$
  - Combine: $dz = V\,dy / K_y\,a\,S\,(y\text{-}y^*)$
    - Packing height (integrate):

$$z = \int_1^2 dz = \frac{V}{K_y a S}\int_{y_1}^{y_2}\frac{dy}{y-y^*}$$

H_{OG} [m] = heat of transfer unit; N_{OG} [-] = number of transfer units

  - Integrate expression for N_{OG} numerically (computer) or graphically (exercise)
  - For dilute mixtures (Analytical Log-mean approximation):

$$N_{OG} = \int_{y_1}^{y_2}\frac{dy}{y-y^*} = \frac{y_1 - y_2}{(y-y^*)_{lm}} \qquad (y-y^*)_{lm} = \frac{(y-y^*)_1 - (y-y^*)_2}{\ln((y-y^*)_1/(y-y^*)_2)}$$

Proof of expression for N_{OG}

- y-y* is the difference between the operating and equilibrium lines and for dilute mixtures it changes linearly with x (and with y) so write[1]:

  y-y* = ky + b
- Book (p. 671) says that this «can be integrated» to get the desired result, but skips the details. Here is a proof.
- Integral of 1/(y-y*)=1/(ky+b) is ln(ky+b)/k = ln(y-y*)/k
- Integrate this from end 2 to end 1, get

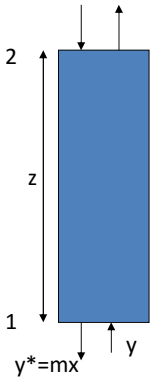
  N_{OG} = ln [$(y-y^*)_1/(y-y^*)_2$] * 1/k
- Now the «trick» comes. Note that k is the slope of (y-y*) as a function of y. So using the two ends 1 and 2, we have

  k = [$(y-y^*)_1$-$(y-y^*)_2$] / [$y_1 - y_2$]

QED

1. Comment: By writing y=(L/V)x+a and y*=mx, we have that y-y*=(L/V - m)x + a, and by using x = (y-a)/(L/V) we find that k = (L/V – m)/(L/V) = 1 – 1/A where A=(L/V)/m

Absorption/stripping in packed towers. Useful log-mean formula for dilute mixtures



Heat exchanger. Assume: constant heat capacities and constant UA

$$Q = mc_p(T_1 - T_2) = UA\Delta T_{lm} \quad [J/s]$$

Countercurrent absorption.

Similar, but instead of $\Delta T = T_h - T_c$, driving force is $\Delta y = y - y^*(x)$ where $y^*(x)$ is in equilibrium with the liquid phase.

Assumptions: 1. Constant slopes for the equilibrium and operating lines (reasonable for dilute mixtures!) 2. Constant mass transfer coefficient ($K_y a$) [mol A / m3].

Then total mass transfer of A from the gas to liquid stream is

$$n_A = V(y_1 - y_2) = K_y \underbrace{(aSz)}_{A}(y - y^*)_{lm} \quad [mol\ A/s]$$

A=aSV [m^2] - total mass transfer area inside the column and

$$(y - y^*)_{lm} = \frac{(y - y^*)_1 - (y - y^*)_2}{\ln((y - y^*)_1/(y - y^*)_2)}$$

1=one end of column, 2=other end

Multistage separation with given products:

Minimum number of stages

- There is no “minimum number of stages” in absorption/stripping
- Can have N -> 0 if we have very large flow (of solvent or stripping gas).
- Note: Purities of incoming streams (column ends) are fixed, and this limits product purities, independent of the number of stages.
- Distillation is different: Here we “recycle” at the column ends and can achieve any product purity by increasing the number of stages.
  - Fenske: $N_{min} = \ln S / \ln \alpha$

Multistage separation with given products:

Minimum flows (Lmin, Vmin)

- Absorption/stripping: minimum solvent/stripping gas
- Obtained with infinite large column (infinite no. of stages)
- General: Occurs when we have “pinch” somewhere

  No change in composition between stages

  = Crossing of equilibrium and operating line

  = Compositions between stages in equilibrium
- Absorption/stripping: usually occurs at column end

  -> one product is in equilibrium with its feed

10.6C Pressure drop and flooding in packed columns

To find column diameter and pressure drop:

1. Obtain packing factor F_p *(from Table)*
2. Find flooding pressure drop

$$\Delta P_{flood}[in.H2O/ft.packing] = 0.115F_p^{0.7} \qquad F_p[ft^{-1}]\text{: packing factor}$$

3. Find corresponding flooding gas velocity *(using pressure drop correlation in Figure)*
   - Typical: v_{flood} is about 6.6 ft/s = 2 m/s
4. Choose diameter such that gas velocity is about half of this
   - Typical: Design for v = 1 m/s
5. Find pressure drop in column *(from pressure drop correlation in Figure)*

Table 10.6-1.
Packing factors F_p

Type		Material	Nominal size, in.	Void fraction, ε	Surface area, a, ft^2/ft^3 (m^2/m^3)	Packing factor, F_p, ft^{-1} (m^{-1})	Relative mass-transfer coefficient, f_p
Random Packing							
Raschig Rings		Ceramic	1/2	0.64	111 (364)	580 (1900)	1.52
			1	0.74	58 (190)	179 (587)	1.20
			1 1/2	0.73	37 (121)	95 (312)	1.00
			2	0.74	28 (92)	65 (213)	0.85
Berl Saddles		Ceramic	1/2	0.62	142 (466)	240 (787)	1.58
			1	0.68	76 (249)	110 (361)	1.36
			2		32 (105)	45 (148)	
Pall Rings		Metal	1	0.94	63 (207)	56 (184)	1.61
			1 1/2	0.95	39 (128)	40 (131)	1.34
			2	0.96	31 (102)	27 (89)	1.14
Metal Intalox (IMTP)		Metal	1	0.97	70 (230)	41 (134)	1.78
			2	0.98	30 (98)	18 (59)	1.27
Nor-Pac		Plastic	1	0.92	55 (180)	25 (82)	
			2	0.94	31 (102)	12 (39)	
Hy-Pak		Metal	1	0.96	54 (177)	45 (148)	1.51
			2	0.97	29 (95)	26 (85)	1.07
		Plastic	1	0.92	55 (180)	25 (82)	
			2	0.94	31 (102)	12 (39)	
Structured Packing							
Mellapak	250Y	Metal		0.95	76 (249)	20 (66)	
	500Y				152 (499)	34 (112)	
Flexipac	2			0.93	68 (223)	22 (72)	
	4			0.98		6 (20)	
Gempak	2A			0.93	67 (220)	16 (52)	
	4A			0.91	138 (452)	32 (105)	
Norton Intalox	2T			0.97	65 (213)	17 (56)	1.98
	3T			0.97	54 (177)	13 (43)	1.94
Montz	B300				91 (299)	33 (108)	
Sulzer	CY	Wire Mesh		0.85	213 (700)	70 (230)	
	BX			0.90	150 (492)	21 (69)	

Data from Ref. (K1, L2, P2, S4). The relative mass-transfer coefficient, f_p, is discussed in Section 10.8B.

$$\Delta P_{flood}[in.H2O/ft.packing] = 0.115F_p^{0.7} \qquad F_p[ft^{-1}]\text{: packing factor}$$
$$\Delta P_{flood}[bar/m\ packing] = 0.00958F_p^{0.7} \qquad F_p[ft^{-1}]\text{: packing factor}$$

- Pressure drop correlations

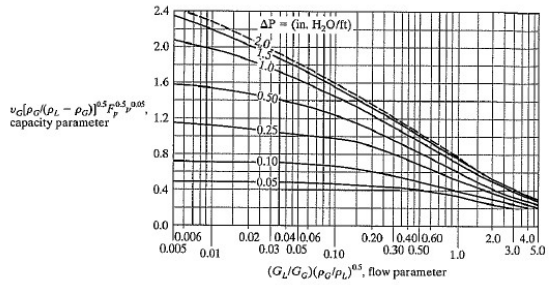


FIGURE 10.6-5. *Pressure-drop correlation for random packings by Strigle. (From R. F. Strigle, Jr., Random Packings and Packed Towers, Houston: Gulf Publishing Company, 1987. With permission from Elsevier Science.)*

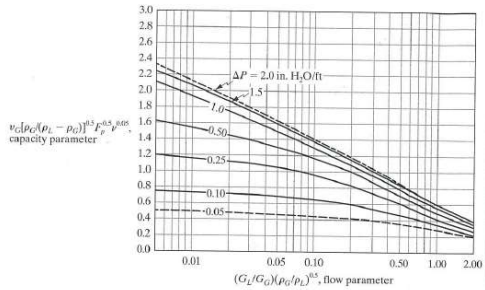


FIGURE 10.6-6. *Pressure-drop correlation for structured packings by Kister and Gill (K2). (From H. Z. Kister, Distillation Design, New York: McGraw-Hill Book Company, 1992. With permission.)*

11.5F Flooding velocity and Diameter of tray towers

To find column diameter and pressure drop:

1. Obtain tray factor K_v [ft/s] *(from Figure)*
2. Find entrainment gas velocity from eq. 11.5-14 (occurs just before flooding). Fair correlation:

$$v_{max} = K_v\left(\frac{\sigma}{20}\right)^{0.2}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \qquad \sigma \approx 20 \text{ dyn/cm for organic liquids, } =72 \text{ for water}$$

$$\text{Typical value: } v_{max} = 3 \text{ m/s}$$

3. Choose diameter so that gas velocity is about $v = 0.7\, v_{max}$
4. Pressure drop: Usually much larger than in packed columns, because of pressure drop for gas to pass through liquid on trays

$$\Delta p = \Delta p_{dry} + \rho_L g h$$

$$\Delta p_{dry} = \text{pressure drop through holes of trays}$$

$$h = \text{sum of liquid levels on all trays (about } 10\% \text{ of column height)}$$

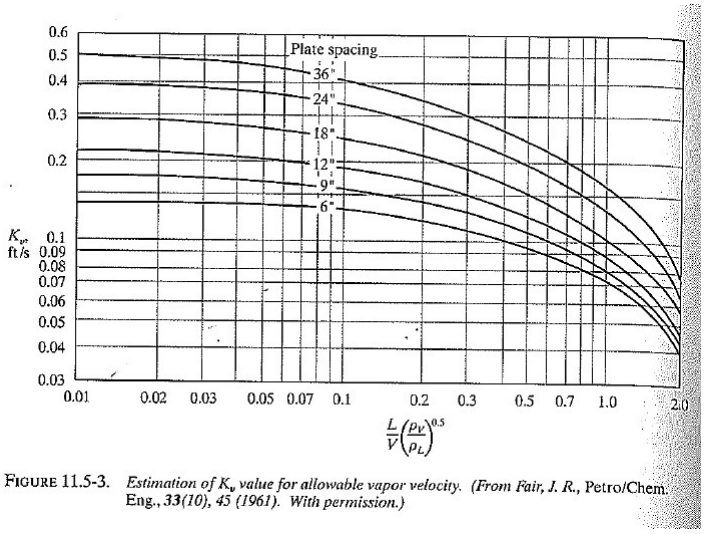


FIGURE 11.5-3. *Estimation of K_v value for allowable vapor velocity. (From Fair, J. R., Petro/Chem. Eng., 33(10), 45 (1961). With permission.)*

6'' = 6 inches = 18 cm

Summary: Absorption/Stripping

- Single section
- Notation varies…
- Dilute mixtures:
  - 1-x ≈1
  - Henry's law: $p_i = H\, x_i$
    - Using $p_i = y_i\, p$ gives
    - **$y_i = (H/p)\, x_i = m\, x_i$**
  - Constant (molar) flows: L ≈ L' (inert)
- Modelling. Two approaches
  1. Equilibrium stage (as for distillation, but simpler)
     - (A) Graphical: McCabe Thiele for nonideal VLE
     - (B) Analytical for dilute mixtures: Kremser formulas. A = (L/V)/m
  2. Mass transfer (Non-equilibrium) models

$$dn_A = K_y a S \cdot (y - y^*) dz \quad (*)$$

1) Introduce $dn_A = V dy$, rearrange and integrate:

$$z = H_{OG} N_{OG}$$

where

$H_{OG} = \frac{V}{K_y a S}$ = height of transfer unit

$N_{OG} = \int_{y_1}^{y_2} \frac{dy}{y - y^*}$ = no.of transfer units

2) Integrate (*) for straight equilibrium and operating lines:

$$\underbrace{n_A}_{V(y_1 - y_2)} = K_y a S z \cdot (y - y^*)_{lm}$$

$$\Rightarrow N_{OG} = \frac{y_1 - y_2}{(y - y^*)_{lm}}$$

Summary countercurrent vapor-liquid separation

	Stage model (Tray column)	Differential model (Packed column)
Equilibrium (between liquid and vapor bulk phases)	**Eq. stage. (Most common model in practice!)** **McCabe-Thiele** **Dilute: Fenske, Kremser**	(Not possible with differential model*) **(But one can use eq. stage + HETP!)**
Non-equilibrium	(Not covered in this course) Non-equilibrium between bulk phases is used sometimes. May use two-film theory on stages	**Most common is two-film approach.** **Nondilute: Numerical integration** **Dilute mixtures: log-mean formula.**

*Would give column with height=0

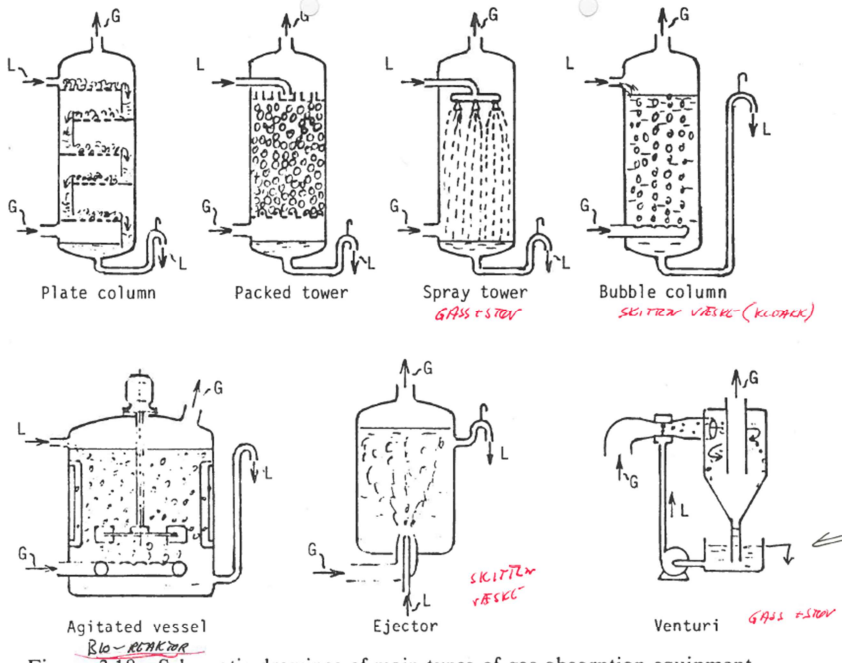


Figure 3.18 Schematic drawings of main types of gas absorption equipment