NTNU – Trondheim
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Department of chemical engineering

Examination paper for TKP4105 (Separation Technology)

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permitted. Simple calculator code D

accepted.

Other information:

Language: English

Number of pages: 5 (frontpage included)

Number of pages attachment enclosed: 2

mm-paper is needed!

Checked by:

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Date   Signature

Problem 1 Gas separation with membranes 25%

A feed stream with mixed gas contains 50% of gas A and 50% of gas B. These gases are to be separated with a membrane. Gas A is the fastest permeating gas, while gas B is the valuable gas which we want to be as clean as possible. (This mixed gas can for instance be biogas, in other words, A = CO_2 and B = CH_4)

Use ideal mixing model for the calculations. Useful information about the system is given below, and also equations in the Attachment 1.

a) Calculate the minimum concentration of component A which can be achieved in the retentate stream. How clean is in this case gas B?
b) Assume that we are, in our case, able to achieve a concentration of A in the retentate which is 25% higher than the one you have calculated in a). What will then be the composition of the permeate stream? How much of gas B (give as vol%) will be lost in the permeate stream?
c) How large membrane area is needed to perform this separation?
d) Do you have any suggestions on how to reduce the loss of B by doing some changes to the process?

Gitt for systemet:

Feed stream, q_f = 2 m^3(STP) / min
Feed pressure: 10 bar
Permeate pressure: 1 bar   (1 bar = 76 cm Hg)
Membrane thickness: t = 10^{-6} m
Permeability of A: P'_A = 400 Barrer   (1 Barrer = 10^{-10} cm^3(STP)·cm/(s·cm^2·cm Hg)
Selectivity of P'_A/P'_B: α = 30
Fraction of the feed permeated: θ= 0,53
*See also Attachment 1 for equations*

Problem 2 Distillation   25%

A distillation column has 7 theoretical stages (reboiler + 3 ideal plates below the feed + 2 ideal plates over the feed + partial condenser). The feed is 10 kmol/h and is a binary saturated liquid with x_F=0.5 (mole fraction A). The column operates at 2 atm, and the xy equilibrium data at 2 atm is given in the figure.

Operation specifications: The top product (D) should have y_D=0.90 (mole fraction A) and the molar liquid to vapor ratio in the upper part of the column should be L/V=0.86.

(a) (i) Find the mole fraction x_B of component A in the bottom product and find the amount of desitiilate D and bottom product B [kmol(h].
(ii) How does the answer change if the feed is moved one stage down (so that there are 3 ideal plates above the feed)?

(b) Control (can be done independently): Suggest a control structure which satisfies the oparation specifications. (What are the MVs, CVs and suggested pairings?)

(See next page for 2 figures given)
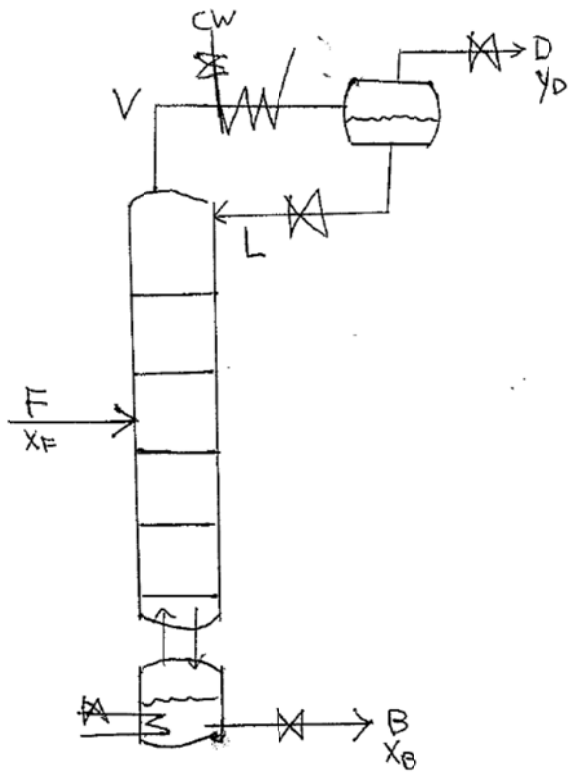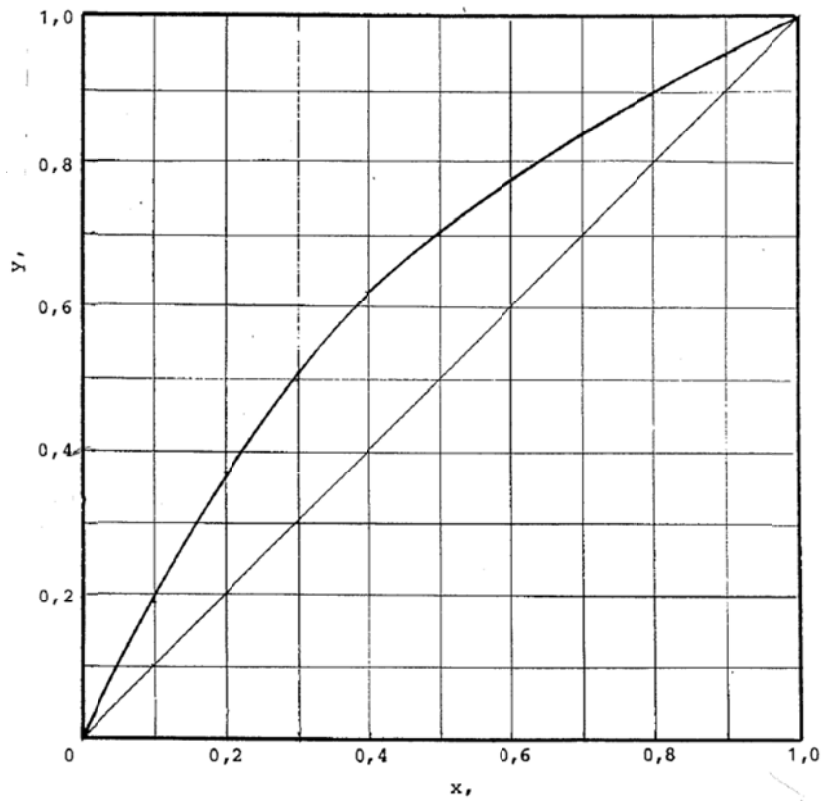
Figures for Problem 2

Problem 3 - Absorption 25%

V_2 $\uparrow$ L_2
Y_2 $\downarrow$ X_2

$$\frac{Y_1 - Y_1^*}{Y_2 - Y_2^*} = A^N \quad (1)$$

$$A = \frac{L/V}{m} \quad (2)$$

$$A = \frac{Y_1 - Y_2}{Y_1^* - Y_2^*} \quad (3)$$

V_1 $\uparrow$ L_1
Y_1 $\downarrow$ X_1

The flue gas (V_1) from a coal power plant is 3000 mol/s (84 kg/s) at 1bar and contains 0.2 mol% SO2 (y_1=0.002). We want to remove at least 95% of the SO2 (y_2<0.0001) by absorption with 90000 mol/s (1600 kg/s) of sea water (x_2=0) at 25 C (wet scrubbing); see the Figure.

Equilibrium data (Henry's constant for SO2 in water; p_{SO2} = Hx where x is mole fraction of SO2):
H=115 bar (50°C), H=46 bar (25°C), H=22 bar (5°C).

(a) Even with a large column (with a large number of equilibrium stages, N) one was not able to reach the specification. Can you explain why? What is the SO2 contents (y_2) in the exit gas if you assume a very large column (N $\to\infty$)?

(b) To remove more SO2, it is suggested build an additional column, that is, the exit gas (y_2) is send to a second identical absorption column (which also has 90000 mol/s of sea water and a large number of stages). Make a flowsheet of the proposed process. Does this help? What is the SO2 contents (y_2') after the second absorption column with a large column (N $\to\infty$)?

(c) Instead of building a second column, the following modifications for the original design have been suggested:
   a. Increase the sea water flow by a factor 2
   b. Decrease the temperature from 25°C to 5°C.
   c. Increase the pressure from 1 bar to 2 bar

   Would any of these modifications give acceptable SO2 removal? Which modification would you suggest?

(d) For calculations in part (c) you may use the Kremser equation (1) given in the figure with N=10.
   a. What are the assumptions behind equation (1)?
   b. How are y_1^* and y_2^* defined?
   c. A is defined by (2), but it is also given by (3). Use an xy-diagram with operation line and equilibrium line (McCabe-Thiele) to prove that (3) is correct.

Problem 2   Adsorption 25%

Using molecular sieves, water vapor is removed from N2 gas in a packed bed column. At 28°C. The column height is 0.3 m and has a bulk density of the solid equal to 712.8 kg/m3. The initial water concentration in the solid is 0.01 kg H_2O / kg solid, and the mass velocity of the N_2 gas is 4052 kg/(m^2·h). The initial water concentration coming in with the gas is c_0 = 926x10^{-6} kg H_2O / kg N_2

The break through data for mass transfer zone is given below. A value of c/c_0 = 0.02 is preferred for the break point.

t (h)	0	9	9.2	9.6	10	10.4
c (kg H_2O/kg N_2 × 10^6)	<0.6	0.6	2.6	21	91	235
t (h)	10.8	11.25	11.5	12.0	12.5	12.8
c (kg H_2O/kg N_2×10^6)	418	630	717	855	906	926

a) Draw the curve c/c_0 as function
b) Determine the break point time, and the height of "unused bed"
c) Determine the fraction of total capacity used up to this point.
d) Determine the loading capacity of the column for uptake of water up to the break point.

3.4 COMPLETE-MIXING MODEL FOR GAS SEPARATION BY MEMBRANES

3.4A Basic Equations Used

n Fig. 13.4-1 a detailed process flow diagram is shown for complete mixing. When a eparator element is operated at a low recovery (i.e., where the permeate flow rate is a small fraction of the entering feed rate), there is a minimal change in composition. Then the results derived using the complete-mixing model provide reasonable esti- nates of permeate purity. This case was derived by Weller and Steiner (W4).

The overall material balance (Fig. 13.4-1) is as follows:

$$q_f = q_o + q_p \qquad (13.4\text{-}1)$$

vhere q_f is total feed flow rate in cm^3(STP)/s; q_o is outlet reject flow rate, cm^3(STP)/s; ind q_p is outlet permeate flow rate, cm^3(STP)/s. The cut or fraction of feed permeated,), is given as

$$\theta = \frac{q_p}{q_f} \qquad (13.4\text{-}2)$$

The rate of diffusion or permeation of species A (in a binary of A and B) is given below by an equation similar to Eq. (6.5-8) but which uses cm^3(STP)/s as rate of permeation rather than flux in kg mol/s · cm^2.

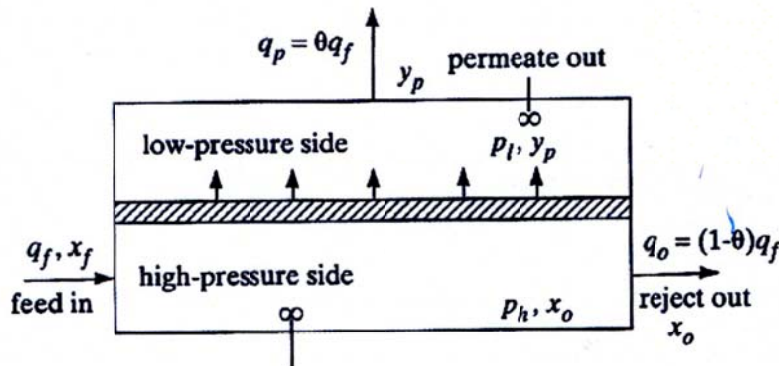$$\frac{q_A}{A_m} = \frac{q_p y_p}{A_m} = \left(\frac{P'_A}{t}\right)(p_h x_o - p_l y_p) \qquad (13.4\text{-}3)$$

FIGURE 13.4-1. *Process flow for complete mixing case.*

Chap. 13   *Membrane Separation Processes*

where P'_A is permeability of A in the membrane, cm³(STP)·cm/(s·cm²·cm Hg); q_A is flow rate of A in permeate, cm³(STP)/s; A_m is membrane area, cm²; t is membrane thickness, cm; p_h is total pressure in the high-pressure (feed) side, cm Hg; p_l is total pressure in the low-pressure or permeate side, cm Hg; x_o is mole fraction of A in reject side; x_f is mole fraction of A in feed; and y_p is mole fraction of A in permeate. Note that $p_h x_o$ is the partial pressure of A in the reject gas phase.

A similar equation can be written for component B.

$$\frac{q_B}{A_m} = \frac{q_p(1 - y_p)}{A_m} = \left(\frac{P'_B}{t}\right)[p_h(1 - x_o) - p_l(1 - y_p)] \tag{13.4-4}$$

where P'_B is permeability of B, cm³(STP)·cm/(s·cm²·cm Hg). Dividing Eq. (13.4-3) by (13.4-4)

$$\frac{y_p}{1 - y_p} = \frac{\alpha^*[x_o - (p_l/p_h)y_p]}{(1 - x_o) - (p_l/p_h)(1 - y_p)} \tag{13.4-5}$$

This equation relates y_p, the permeate composition, to x_o, the reject composition, and the ideal separation factor α^* is defined as

$$\alpha^* = \frac{P'_A}{P'_B} \tag{13.4-6}$$

Making an overall material balance on component A

$$q_f x_f = q_o x_o + q_p y_p \tag{13.4-7}$$

Dividing by q_f and solving for the outlet reject composition,

$$x_o = \frac{x_f - \theta y_p}{(1 - \theta)} \quad \text{or} \quad y_p = \frac{x_f - x_o(1 - \theta)}{\theta} \tag{13.4-8}$$

Substituting $q_p = \theta q_f$ from Eq. (13.4-2) into Eq. (13.4-3) and solving for the membrane area, A_m,

$$A_m = \frac{\theta q_f y_p}{(P'_A/t)(p_h x_o - p_l y_p)} \tag{13.4-9}$$

13.4C Minimum Concentration of Reject Stream

If all of the feed is permeated, then $\theta = 1$ and the feed composition $x_f = y_p$. For all values of $\theta < 1$, the permeate composition $y_p > x_f$ (H1). Substituting the value $x_f = y_p$ into Eq. (13.4-5) and solving, the minimum reject composition x_{oM} for a given x_f value is obtained as

$$x_{oM} = \frac{x_f\left[1 + (\alpha^* - 1)\dfrac{p_l}{p_h}(1 - x_f)\right]}{\alpha^*(1 - x_f) + x_f} \tag{13.4-12}$$

Hence, a feed of x_f concentration cannot be stripped lower than a value of x_{oM} even with an infinitely large membrane area for a completely mixed system. To strip beyond this limiting value a cascade-type system could be used. However, a single unit could be used which is not completely mixed but is designed for plug flow.