SOLUTION Konte-eksamen 10. august 2010. Problems 1-4.

Oppgave / Oppgåve 1: Ett-trinns flash (20%)

(a) Den relative flyktigheten mellom komponentene 1-2 er definert som $\alpha = (y_1/x_1) / (y_2/x_2)$.

(a1) Bruke dataene under til å bestemme α for propan-butan (bruk α = 6 hvis du ikke får det til).

(a2) Utled sammenhengen mellom y_1 og x_1 (molfraksjon lett komponent) for en binær blanding med konstant relative flyktighet.

(b) Tegn xy-diagrammet og skisser Txy-diagrammet for propan-butan ved 1 atm.

(c) Flash. Følgende to strømmer blandes:

Strøm F_a: 80 mol/s. Mettet damp ved 1 atm. 50% propan. 50% n-butan

Strøm F_b: 70 mol/s. Mettet væske ved 1 atm. 50% propan. 50% n-butan

Den blandede strømmen separates in adiabatisk flashtank ved 1 atm og gir er væskeprodukt L and gassprodukt V (i likevekt). Bestem mengde og sammensetning av strømmene L og V.

- Anta konstant relativ flyktighet α.
- *Energibalanse:* Anta konstante molare strømmer.

(d) Gi en rangering (fra høy til lav) av temperaturene i strømmene F_a, F_b, L og V (grunngi svaret ditt).

Damp/væske-likeveltsdata:

	Kokepunkt (1 atm)	Damptrykk ved 240K
Propan (1)	231.14K	1109 mmHg
Butan (2)	272.98K	180.6 mmHg

Oppgave / Oppgåve 2. Destillasjon (10%)

En væskeføde med to komponenter (50% A og 50%B) med relativ flyktighet 6 skal separeres ved kontinuerlig destillasjon i et topp-produkt og bunnprodukt. Hva er minimum antall teoretiske trinn som trenges for å få produkter 0.1% A (bunn) og 1% B (topp). Hvor mange teoretiske trinn (omtrentlig) vil du anbefale i en virkelig kolonne?

Oppgave / Oppgåve 3. Formel (10%)

Gitt formelen

$$z = \frac{V}{K_y a S} \int_{y_{A1}}^{y_{A2}} \frac{dy}{y_{AG} - y_A^*}$$

(a) Definer alle variablene og gi enheter for dem. Lag et flytskjema som illustrerer betydningen av variablene

(b) Hva brukes formelen for og hvilke antagelser er den basert på? Hvordan kan den løses?

(c) Hva er N_A uttrykt ved variablene i formelen? Hva er HTU, NTU, H_{OG} og N_{OG}?

Oppgave / Oppgåve 4. Regulering (10%)

Tegn et flytskjema av en kontinuerlig destillasjonskolonne. Varmetilførselen (oppkok V) er på sin maksimalverdi (for å få til best mulig separasjon). Foreslå en reguleringsstruktur basert på tilbakekobling (med to LC'er, en PC og en TC). Fødestrømmen er gitt ("forstyrrelse").

Problem 1: Single-stage flash (20%)

(a) The relative volatility between components 1-2 is defined as $\alpha = (y_1/x_1) / (y_2/x_2)$.

(a1) Use the data below to find α for propane-butane (use $\alpha = 6$ if you cannot find it).

(a2) Derive the relationship between y_1 and x_1 (mole fractions light component) for a binary mixture with constant relative volatility.

(b) Sketch the equilibrium Txy-diagram and xy-diagram for the system propane-butane at 1 atm.

(c) Flash. The following two streams are mixed

Stream F_a: 80 mol/s. Saturated vapour at 1 atm. 50% propane. 50% n-butane

Stream F_b: 70 mol/s. Saturated liquid at 1 atm. 50% propane. 50% n-butane

The mixed stream is separated in an adiabatic flash unit operating at 1 atm into a liquid product L and vapour product V (in equilibrium). Find the amount and compositions of the streams L and V.

- Assume constant relative volatility α.
- *Energy balance:* Assume constant molar flows.

(d) Rank the temperatures of the streams F_a, F_b, L and V (justify your answer).

Vapor-liquid equilibrium data:

	Boiling point (1 atm)	Vapor pressure at 240K
Propane (1)	231.14K	1109 mmHg
Butane (2)	272.98K	180.6 mmHg

Problem 2: Distillation (10%)

Consider the separation of a liquid feed with two components (50% A and 50%B) with a relative volatility of 6 into a distillate and bottoms product. What is the minimum number of distillation stages required to get products with 0.1% A (bottoms) and 1% B (top). What would you recommend as the actual number of stages (approximately)?

Problem 3. Formula (10%)

Consider the formula

$$z = \frac{V}{K_y a S} \int_{y_{A1}}^{y_{A2}} \frac{dy}{y_{AG} - y_A^*}$$

(a) Define all the variables, including giving their units. Make a flowsheet with the variables.
(b) What is this formula used for and what assumptions are required? How can it be solved?

(c) What is N_A given in terms of the variables in the formula? What is HTU, NTU, H_{OG} and N_{OG}?

Problem 4. Control (10%)

Make a flowsheet of continuous distillation column. The heat input (boilup V) is set at its maximum (to maximize separation). Suggest a feedback control structure (involving two LC's, one PC and one TC). The feed stream to the column is given ("disturbance").

SOLUTION Konte-eksamen 10. aug. 2010

Solution Problem 1

(a1) From the vapor pressures at 240 K we have, $\alpha = p_1/p_2 = 1109/180.6 = 6.14$.

Extra: Check using Sigurd's formula based on boiling points.
At normal boiling point dHvap is 18.8 kJ/mol and 22.4; average is 20.6 kJ/mol.
Geometric avg. boiling point is Tb = sqrt(231.14*272.98)=251.2 K. So dHvap/RTb=9.87.
Boling point difference = 272.98 - 231.14 = 41.84 K
Estimate alpha = exp(dHvap/RTb * dTb/Tb) = exp (9.87*41.84/251.2) = 5.18., which is a bit low.

(a2) For a binary mixture of components 1 and 2, the relative volatility is defined as:

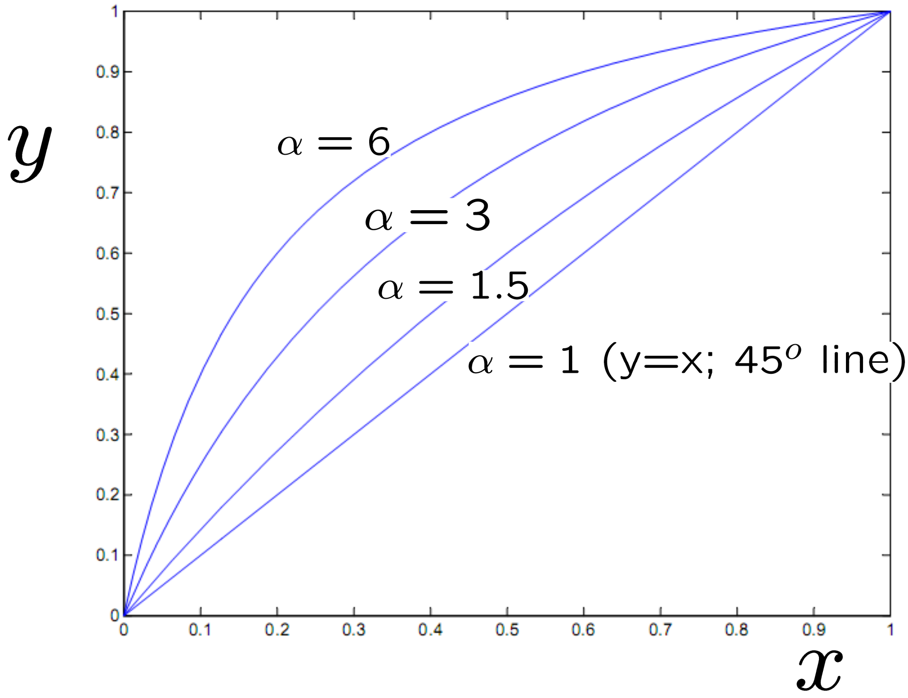
$\alpha = (y_1/x_1) / (y_2/x_2) = (y_1/x_1) / ((1-y_1)/(1-x_1))$

Solve with respect to y_1:
$y_1 = \alpha x_1 / (1 + (\alpha-1)x1)$

Drop in the following index 1 for the light component:
$y = \alpha x / (1 + (\alpha-1)x)$

(b) The Txy diagram is similar to Figure 11.1-1 with temperatures starting at 272.98K and ending at 231.14 K. The saturated vapour and liquid lines will be further apart than in Figure 11.1.-1 because of the high relative volatility.

The xy-diagram is similar to Figure 11.2.-1 but the curve is higher up because of the large relative volatility.



(c) Flash. Mixing:

Overall feed: F = Fa + Fb = 150,
xF = 0.5

Flash: Overall mass balance

F = V + L

Flash: Mass balance light component (1)

xF F = y V + x L

Constant molar flow assumption: L = 70, V=80.

VLE using alpha=6 (alternatively, use alpha=6.14): y = 6x /(1+5x)

Get: 75 = 80*6x/(1+5x) + 70*x
350 x^2 + 175 x – 75 = 0

x = 0.2761
y = 0.6959

Check mass balance

y V + x L = 0.2761*70 + 0.6959*80 = 75 (OK!)

Comment: A graphical solution is also possible for this problem; see Fig. 11.2-1 (make q-line with slope L/V).

(d) The rank is
T_{Fa} (hottest) > T_L=T_V (same since they are in equilibrium) > T_{Fb} (coldest)

Why? The liquid in equilibrium with Fa has more than 50% butane since the heavy component accumulates in the liquid. It is therefore "heavier" than Fb which only has 50% butane, and will therefore boil at a higher temperature

Solution Problem 2

With constant relative volatility and total reflux we derive the Fenske equation (all mole fractions for light component)

$S = (x_D/(1-x_D)) / (x_B/(1-x_B)) = \alpha^{Nmin}$

Nmin = ln S / ln α

$S = (x_D/(1-x_D)) / (x_B/(1-x_B))$ = (0.99 /0.01)/ (0.001/0.999) = 98901

Minimum number of theoretical stages (including reboiler):

Nmin = 11.50/ 1.79 = 6.42

Comment: Can also use alternative solution with McCabe Thiele steps, but difficult to plot with so high relative volatility.
Note that for Nmin it does not matter if the feed is liquid or vapour.

To use this number of stages we would need total reflux (infinite energy).
In a real column we should use between 2 and 3 times more stages, that is, should use N about 15.

Solution Problem 3. Absorption (10%)

Consider the formula

$$z = \frac{V}{K_y aS} \int_{y_{A1}}^{y_{A2}} \frac{dy_A}{y_{AG} - y_A^*}$$

(a) Define all variables

z – height of column [m]
V – vapour flow [mol/s]
K_y – overall gas phase transfer coefficient [mol/s, m2 mass transfer area]
a – specific area [m2 mass transfer area / m3 packing]
S – cross-sectional area of column [m2]
y_A – gas mole fraction of component A
y_{AG} – .. in bulk gas phase
y^*_A - … in equilibrium with bulk liquid phase (x_{AL})
Note: Volume of column [m3] = z S

Flowsheet: See Fig. 10.6-9

(b) The formula is used to compute the required packing height in an absorption (or stripping) column.
Assume constant V, K_y, a, S along the column.

Derivation (not asked for): Set up a differential mass balance over a column section dz[m]:
$dN_A = d(Vy) = d(Lx)$
where from film theory approach with diffusion mass transfer between phases,
dN_A[mol/s] = N_A dA where
N_A[m2 mass transfer area]=aSdz and N_A=K_y (yAG – yA*) [mol, s m2 mass transfer area].
Using $dN_A = d(Vy)$ gives desired formula.
Note: The “mass balance” part d(Vy) = d(Lx) gives the “operating line” which is needed for the intergration of the fomula.

Solution: Must also use VLE= equilibrium line (gives yA* is a function of x_{AL}) and mass balance=operating line (gives x_{AL} as a function of y_{AG}).
Can then solve numerically or graphically.
Can be solved analytically in some cases. For example, assuming Henrys law ($y_A^*=Hx_{AL}$), gives that the integral is equal to the mean-log difference, $1/(y_{AG}-y_A^*)_{LM}$.

(c) Local N_A = K_y (yAG – yA*) [mol, s m2 mass transfer area]
HTU [m] = $H_{OG} = V/ K_y aS$

$$\text{NTU} = \text{N}_{\text{OG}} = \int_{y_{A1}}^{y_{A2}} \frac{dy_A}{y_{AG} - y_A^*}$$

Problem 4. Control (10%)

- Reboiler dury (QB, V) is at max. and is not available for control.
- Feed (F) is a disurbance

Proposed control structure:

1. LC: Top level (Md) controlled using top product (D).
2. LC: Btm level (Mb) controlled using btm product (B)
3. PC: pressure controlled using cooling (QD)
4. TC: Sensitive temperature in top (T) controlled using reflux (L)