

SIEVE TRAY DISTILLATION COLUMN

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1. GENERAL

Only continuous distillation (or stationary distillation) is considered here. Batch distillation, which is dependent of time, is not involved in this work. Distillation as a continuous and industrial unit operation occurs usually in a single device, which is called a distillation column. Sometimes several columns are used and they can be connected to each other in many ways.

Distillation as a unit operation means separating components from each other. Separation is based on different vapor pressures of components. In figure 1 there is a simplified scheme of distillation process in one column.

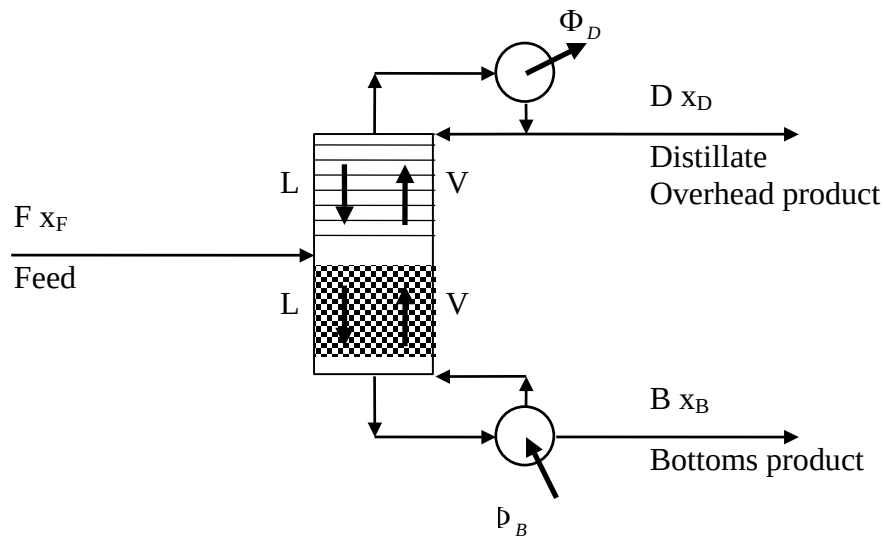


Figure 1. Simplified scheme of distillation process.

In practice continuous distillation occurs in distillation columns where vapor phase V vaporized in a reboiler by adding a heat flux Φ_B goes upwards and liquid phase L (from condenser by subtracting heat flux Φ_D) goes downwards. The column is not empty, there are usually either distillation trays or packing in order to increase the contact area between the two phases. In figure 1 there is an unlikely case where the enriching section is a tray column and stripping section is a packed column.

Phases on the same level in a column have different concentrations, which differ from the equilibrium concentration and therefore the components transfer between the two phases (actually due to their chemical potentials).

2. THEORY

A mixture of n-hexane and cyclohexane is distilled in this work. The mixture of these two components is very ideal and later it will be marked that the mass transfer is quite equimolar. Due to this the flow rates are almost constant and this case is comparable to the one, which is discussed in McCabe (pages 529 – 540).

In order to calculate the amount of ideal steps in this ideal case, the phase equilibrium curve, the feed and product flows, and the inner flows in the column must be known. The concentrations of the feed and product flows are measured and the inner flows are calculated from the balances.

2.1 NOTATIONS

The behavior of the lightest component, in other words the component that boils at lowest temperature and has the highest vapor pressure, is studied (especially in binary case). This component is usually marked with A, but the subscript is left out: concentration that has no subscript is component A and $x_F = x_{A,F}$.

From this choice follows that $x_D > x_B$ and the distillation column works as a stripper for component A. The operating line is below the equilibrium line and the concentration of component A decreases in the liquid phase when moving downwards the column. The trays are numbered from up to down.

2.2 OPERATING LINES

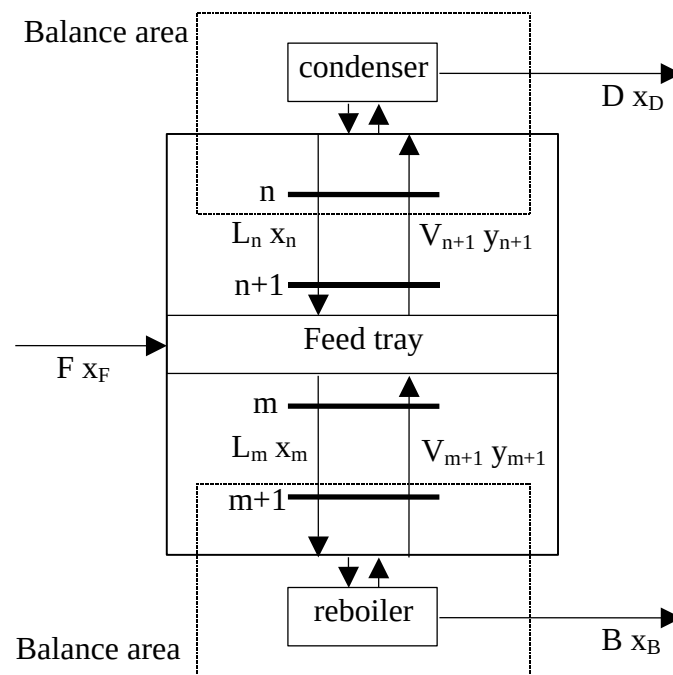


Figure 2. A simple model for distillation column

2.2.1 Enriching Section

The overall balance and the total material balance for component A over the enriching section gives.

$$V_{n+1} = L_n + D \quad (1)$$

$$V_{n+1}y_{n+1} = L_nx_n + Dx_D \quad (2)$$

When dividing equation (2) with the vapor flow V , the equation of the **operating line** in the enriching section is obtained:

$$y_{n+1} = \frac{L_n}{V_{n+1}} x_n + \frac{Dx_D}{V_{n+1}} \quad (3)$$

If the flow rates are constant as here, equation (3) becomes

$$y_{n+1} = \frac{L}{V} x_n + \frac{Dx_D}{V} \quad (4)$$

From the overall balance (1) and equation (4), when divided by the distillation flow D, equation (5) will be obtained:

$$y_{n+1} = \frac{L/D}{L/D+1} x_n + \frac{x_D}{L/D+1} \quad (5)$$

The reflux ratio is defined as

$$R_D \equiv \frac{L}{D} \quad (6)$$

By substituting this into equation (5) another equation for the operating line of enriching section is obtained:

$$y_{n+1} = \frac{R_D}{R_D+1} x_n + \frac{x_D}{R_D+1} \quad (7)$$

When $x_n = x_D$ is substituted to the equation (3, 4 or 7) of the operating line of the enriching section, $y_{n+1} = x_D$ is obtained. This means that the operating line of the enriching section intersects the diagonal ($y = x$) at the point (x_D, x_D) .

2.2.2 Balances of the Condenser

Reflux ratio can be obtained from the balances of the condenser (McCabe, 532-534).

If the condenser is assumed to be an ideal total condenser then

- 1) All vapor condenses in the condenser and
- 2) Only latent heat exits the condenser (liquid is in its boiling point)

Now, the temperatures and concentrations of the fluids V_1 , L_0 , and D are the same (figure 3).

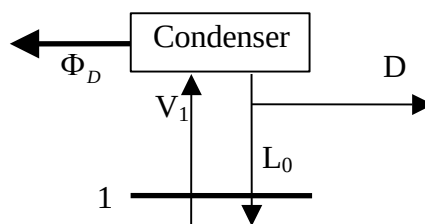


Figure 3. Total condenser

The overall balance and the heat balance over the total condenser are ($i_D = i_0$):

$$\text{Overall balance: } V_1 = L_0 + D \quad (8)$$

$$\text{Heat balance: } V_1 I_1 = L_0 i_0 + D i_0 + \Phi_D = (L_0 + D) i_0 + \Phi_D \quad (9)$$

Solving V from these two equations and equation (10) gives

$$V = \frac{\Phi_D}{I_1 - i_0} = \frac{\Phi_D}{I(T_D) - i(T_D)} = \frac{\Phi_D}{r(T_D)} \quad (10)$$

And finally the reflux ratio is obtained:

$$L = V - D \quad (11)$$

$$R_D = \frac{L}{D} \quad (12)$$

2.2.4 Stripping Section

The overall balance and the total material balance for component A over the stripping section gives.

$$L_m = V_{m+1} + B \quad (13)$$

$$L_m x_m = V_{m+1} y_{m+1} + B x_B \quad (14)$$

By dividing equation (14) with the vapor flow rate V_{m+1} an equation for the operating line of the stripping section is obtained:

$$y_{m+1} = \frac{L_m}{V_{m+1}} x_m - \frac{B x_B}{V_{m+1}} \quad (15)$$

If the flow rates are constant as in this case, equation (15) can be represented as

$$y_{m+1} = \frac{\bar{L}}{\bar{V}} x_m - \frac{B x_B}{\bar{V}} \quad (16)$$

As the reflux ratio in the enriching section, a reboil ratio can be defined similarly to the stripping section. The following is for a case where the flow rates are constant. The reboil ratio is defined by

$$R_B \equiv \frac{\bar{V}}{B} \quad (17)$$

Substituting equations (13) and (17) to equation (16) another form of the operating line in the stripping section is obtained.

$$y_{n+1} = \frac{R_B + 1}{R_B} x_m - \frac{x_B}{R_B} \quad (18)$$

So, equations (15), (16) and (18) all represent the equation of the operating line in the stripping section and the operating line intersects the diagonal on equilibrium curve $y = x$ at the point (x_B, x_B) .

2.2.5 Feed Tray

The reboil ratio is not usually used to calculate the flow rates in the stripping section but the balances of the feed tray.

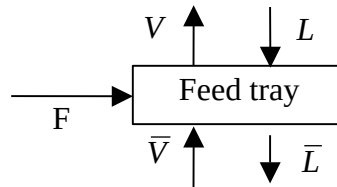


Figure 4. Feed tray.

Now, the following **assumptions** are made:

- 1) There are no pure heat flows (in or out) at the feed tray and
- 2) The enthalpies of L and $\bar{L}$ and the enthalpies of V and $\bar{V}$ are equal.

The latter assumption is reasonable if the concentration gradient over the feed tray is not too significant. Notice that the flows L , $\bar{L}$, V , and $\bar{V}$ are saturated because they are internal flows of the column.

The balances of the feed tray are

$$F + \bar{V} + L = V + \bar{L} \quad (19)$$

$$Fi_F + \bar{V}I + Li = VI + \bar{L}i \quad (20)$$

where i and I are the enthalpies of liquid and vapor on the feed tray, respectively. Combining equations (19) and (20) leads to

$$Fi_F + (\bar{L} - L - F)I = (\bar{L} - L)i. \quad (21)$$

And furthermore

$$\bar{L} - L = \frac{i_F - I}{i - I} F \equiv qF. \quad (22)$$

Finally, when equation (22) is substituted into equation (19), the flow rates can be solved:

$$\bar{L} = L + qF \quad (23)$$

$$\bar{V} = V + (q - 1)F \quad (24)$$

where q describes the state of the feed:

- $q > 1$, for cold liquid feed.
- $q = 1$, for saturated liquid feed.
- $0 < q < 1$, if feed is a mixture of liquid and vapor, and q is the proportion of liquid in the feed.
- $q = 0$, for saturated vapor feed.

By assuming that the feed is cold liquid and by manipulating the definition of q (22)

$$q \equiv \frac{i_F - I}{i - I} = \frac{i_F - i + i - I}{i - I} = 1 + \frac{i - i_F}{I - i} \approx 1 + \frac{\overline{c_{p,F}}(T_{\text{FEED TRAY}} - T_F)}{r(T_{\text{FEED TRAY}})} \quad (25)$$

If the feed is cold liquid as in here, q can be calculated from equation (25) where $\overline{c_{p,F}}$ is the mean heat capacity of the feed in the temperature range $(T_{\text{FEED TRAY}}, T_F)$. Since the mixture is very ideal, the

properties of the mixture OM_{MIX} can be solved from the properties of pure compound OM_c by weighting with the mole fractions (usually the calculations are much complicated).

$$OM_{MIX} = x_A OM_A + x_B OM_B \quad (26)$$

The latent heat $r(T_{FEEDTRAY})$ is the latent heat of the mixture in the temperature and concentrations of the feed tray. When q is calculated, the flow rates are obtained from equations (23) and (24).

2.3 INTERSECTION OF THE OPERATING LINES

The intersection can be solved in two ways:

- From the q -line (feed condition).
- From operating lines.

The q -line is a straight line, which passes the point (x_F, x_F) and the intersection of the two operating lines. The usual procedure when calculating graphically a binary distillation column is

1. Calculate the reflux ratio R_D
2. Draw the operating line of the enriching section (known, when R_D and x_D are known)
3. Draw the q -line (known, when q and x_F are known)
4. Draw the operating line of the stripping section through the intersection of the operating line of the enriching section and the q -line and point (x_B, x_B) .

The intersection of the two operating lines is also obtained from the equations (4) and (16) by demanding that $y_C = y_{n+1} = y_{m+1}$.

$$\frac{L}{V} x_C + \frac{Dx_D}{V} = \frac{\bar{L}}{\bar{V}} x_C - \frac{Bx_B}{\bar{V}} \quad (27)$$

And the intersection x_C can be solved:

$$x_C = \frac{\frac{Dx_D}{V} + \frac{Bx_B}{\bar{V}}}{\frac{\bar{L}}{\bar{V}} - \frac{L}{V}} \quad (28)$$

x_C can be expressed with the help of R_D and q :

$$x_C = \frac{(R_D + 1)x_F - (1 - q)x_D}{R_D + q} \quad (29)$$

3. EQUIPMENT

Column used in this laboratory work is an Oldershaw column; a schema is drawn in figure 5. The structure of the trays in the column is shown in figure 6.

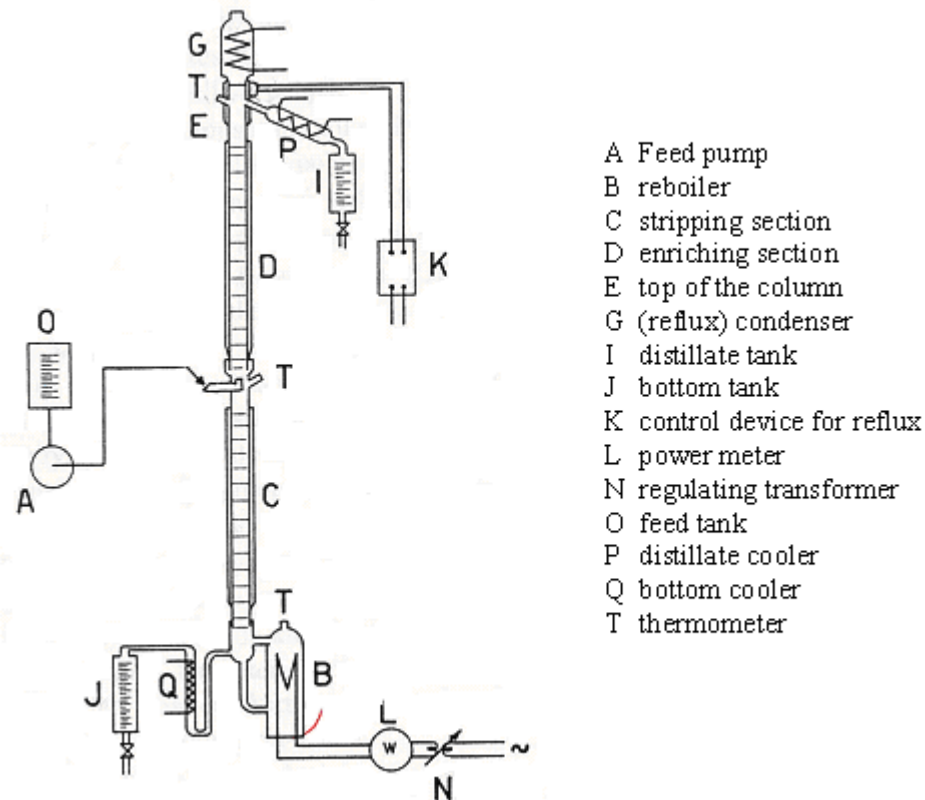


Figure 5. Schematic of Oldershaw Column

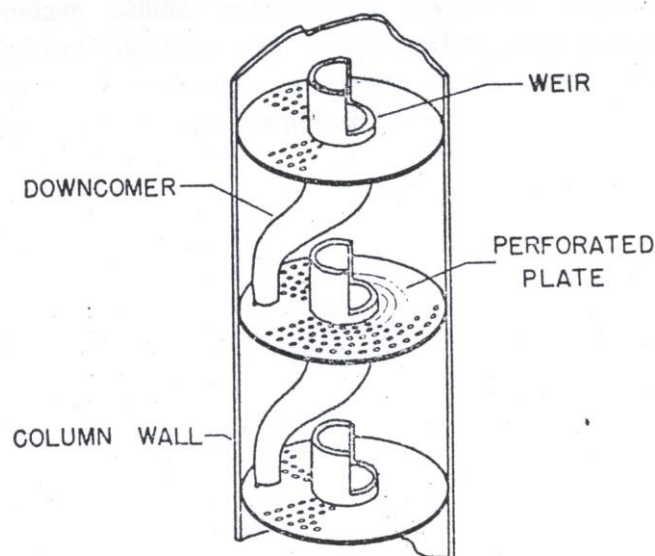


Figure 6. Structure of the column.

3.1 COLUMN

Laboratory column is an Oldershaw sieve tray column, which was first made by C.F. Oldershaw in his laboratory when working in research laboratory of Shell Development Co in California (Oldershaw (1941)). The column and its equipment were delivered by Normschliff Glasgeräte KG in 1965. The column, as the original, is made of glass.

- The height of the stripping section is 0,7 m and there are 10 actual trays.
- The height of the enriching section is 1,1 m and there 20 actual trays.
- The internal diameter of both sections is 47,6 mm and they are covered with silvered vacuum jacket in order to reduce heat losses.
- The trays are sieve trays and in every tray there are 230 holes, diameters of which are 0,95 mm. The distance between two trays is 47,6 mm.
- Reboiler is a natural-circulation - or thermosiphon reboiler. (McCabe, p. 443), and liquid volume is about 1 liter. Heating is done by two heating resistors (quartz rod). Heating power can be adjusted with the regulating transformer and the electrical power can be read from the power meter. Due to heat losses the actual vaporization efficiency is smaller than the measured power.
- Bottom product exits the column bottom as overflow.
- Feed is pumped with a piston pump between the two sections or in practice to the topmost plate of the stripping section, which is the feed plate. The volume flow rate of the feed can be adjusted by adjusting stroke or frequency of the piston pump.
- Condenser is a coil condenser and its heat transfer area is about 0,3 m². The heat power of the condenser can be measured by measuring the flow rate and the temperature change of the cooling water.
- Between the column and the condenser is a reflux control device, which controls the reflux flow rate. There is a pendulum pipe in the control device, which can be adjusted with an electromagnet: When electromagnet is off, all distillate is refluxed and when it is on, all liquid goes to distillate.
- The reflux ratio can be adjusted by changing the impulse intervals of the electromagnet. Due to inaccuracies in this method, the actual reflux ratio is calculated from the heat balance of the condenser.

3.2 MEASURING EQUIPMENT

- stopwatch
- sample bottles
- 2 l measuring cylinder
- measuring tape

4. OPERATING THE DISTILLATION COLUMN

4.1 ARRANGEMENTS

A mixture of n-hexane and cyclohexane is distilled in this work. The physical properties of these components are shown in appendix 1.

CAUTION 1!

The mixture of n-hexane and cyclohexane is very flammable!

CAUTION 2!

Beware of skin contact and inhaling the solvent vapors.

Cyclohexane may irritate skin and perform as a suffocating and narcotic agent. Also it may contain small amounts of benzene (at most 0,05 %).

One **must** keep watch on column when it is in operation.

4.2 STARTING THE WORK

- Open the cooling water valves (2).
- Start the feed pump.
- Assistant gives the boiling power and the reflux ratio.
- Measure the refraction index of the feed (2 times).
- Switch the power on to the reboiler and start heating with 100 W. Notice, that the reading is a quarter of the actual power.
- When vapor reaches the top of the column (temperature!) wait until refluxing starts. The column operates now on total reflux (feed and product flows are zero).
- Switch the reflux control device on and adjust the reflux ratio.
- Adjust the cooling water flow of the condenser so that the change in the temperature of the cooling water is about 5 - 10 K.

4.3 MEASUREMENTS

- After starting the feed measure the refraction index of the distillate and bottom product and the temperatures of the top, bottom, and feed point every 5 minutes.
- When equilibrium is reached, which means that the concentrations and temperatures remains unchanged, make the following measurements:
 1. Take up the boiling power and reflux ratio values.
 2. Measure the feed, distillate, and bottom product mass flow rates simultaneously (at least 200 ml).
 3. Measure refraction indexes of the bottom product and distillate.
 4. Measure the temperatures of the top, bottom, and the feed point of the column.
 5. Measure the cooling water mass flow rate as in 2.
 6. Take up the cooling water temperatures.
 7. Measure the pressure gradient of the column as a difference of two fluid levels.
 8. In order to notice any gross errors, calculate the relative error of the total balance and the n-hexane balance and the heat flow removed from the condenser. Show your results to the assistant.

4.4 FINISHING THE WORK

- Show your results to the assistant and after permission:
 1. Switch the reboiler and feed pump off.
 2. Switch the reflux control device off.
 3. Pour the distillate and bottom product to feed tank.
 4. Leave the cooling water valves open. Assistant closes them when the column is cooled enough.
 5. Clean up your working place.

5. SPECIFICATION

The specification is done as told in general instructions.

Processing of the measuring results:

Show the stabilizing of the column graphically after starting the feed (the concentration and temperature of the top of the column as a function of time).

Change the mass flows to molar flows (all calculations are done in mole fraction coordinates).

Match the measured balances as in appendix 3. The MS-EXCEL spreadsheet can be found in the homepage of this course.

Evaluating the mass transfer ratio:

Assume that

- 1) Heat losses are negligible
- 2) All fluids are saturated
- 3) The enthalpies of saturated vapor and liquid (I and i) are linear functions of the concentration of component A.

With these three assumptions the mass transfer ratio z is constant and it can be calculated from equation (30). Notify the minus signs!

$$z = - \frac{I_B - i_B}{(I_A - I_B) - (i_A - i_B)} = - \frac{I_B - i_B}{(I_A - i_A) - (I_B - i_B)} = + \frac{r_B}{r_B - r_A} \quad (30)$$

Note that since z is very large (> 20), is the mass transfer nearly equimolar ($z = \infty$) and total molar flows are practically constant.

Calculate the pressure loss per a sieve tray.

Summarize the results.

Calculate the reflux ratio from the balances of the condenser ($1 < R_D < 10$).

Calculate q from equation (25).

Determine the amount of ideal stages graphically:

Draw the equilibrium curve and diagonal $y=x$ with MS-EXCEL. According to appendix 1 the relative volatility α_{AB} is constant and $\alpha=1.4$. In a binary case relative volatility is defined as

$$\alpha = \frac{y(1-x)}{x(1-y)} \quad (31)$$

The equation for the equilibrium line can be derived from equation (31):

$$y^* = g(x) = \frac{\alpha x}{1 + (\alpha - 1)x} \quad x^* = g^{-1}(y) = \frac{\frac{1}{\alpha}y}{1 + (\frac{1}{\alpha} - 1)y} = \frac{y}{\alpha + (1 - \alpha)y} \quad (32)$$

Draw the operating line of the enriching section and the q-line. The q-line passes the point (x_F, x_F) and its slope is

$$m = \frac{q}{q - 1} \quad (33)$$

Finally, draw the operating line of the stripping section.

In order to find the number of the theoretical plates, step off the tower. Assume that feed comes to optimum tray (McCabe p 540). Size of the last stage, if it is incomplete, can be evaluated with

$$N = \frac{x_n - x_b}{x_n - x_{n+1}} \quad (34)$$

Calculate the amount of ideal stages in the enriching and stripping sections. The feed plate belongs to stripping section. Assume that reboiler is one ideal stage and subtract one plate from the amount of the ideal plates. Mark the amounts with $N_{G,E}$ and $N_{G,S}$.

Calculate the flow rates of the stripping section from the balances of the feed plate. Calculate the intersection x_c of the operating lines from equation (29).

Determine the amount of ideal stages numerically. The algorithm is:

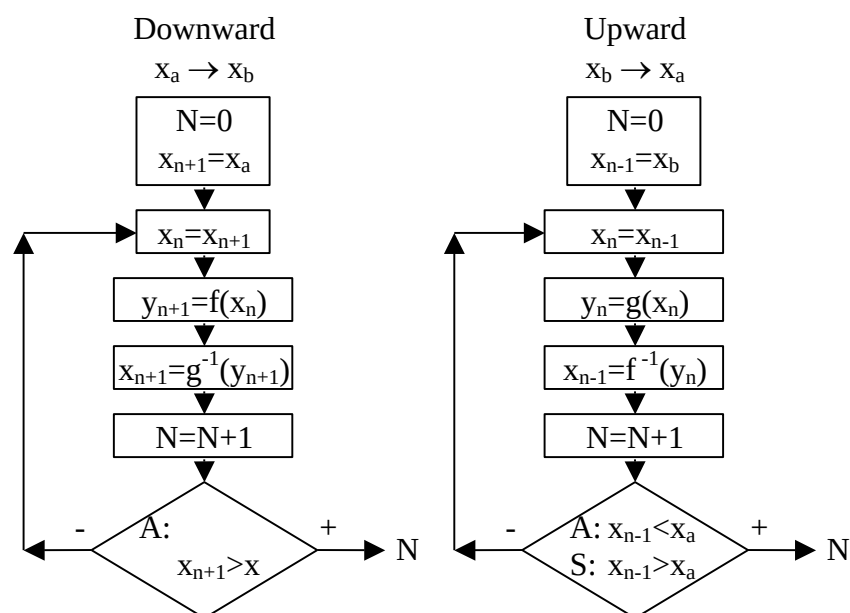


Figure 7. Algorithms for stepping off the tower numerically

Algorithms are valid to both absorbers and strippers. Only difference is the condition for termination of the algorithm, which is different for absorbers (A) and strippers (S). Distillation is analogous to strippers. Distillation column has two sections and two operating lines and therefore stepping is done in two phases. Operating line is changed in the intersection x_C .

Next, a case where numerical stepping is done downward or to the direction of the L-phase. The equations of the operating line (f) and equilibrium line (g) are needed for stepping off the column. The equation of the operating line of the enriching section is derived above and it must be in the form where y_{n+1} can be solved:

$$y_{n+1} = f(x_n) = \frac{L}{V} x_n + \frac{Dx_D}{V} \quad (36)$$

x_{n+1} is calculated in the algorithm so the equation of the equilibrium line must be in the form of:

$$x_{n+1} = g^{-1}(y_{n+1}) = \frac{y}{\alpha + (1 - \alpha)y} \quad (37)$$

Now the enriching section can be stepped off. Initial value is set to $x_{n+1} = x_D$ and the algorithm is continued until $x_{n+1} < x_C$. The obtained N is the value of the ideal stages in the enriching section. The last concentration is marked with x_{N+1} .

The stripping section is done similarly from initial value $x_{n+1} = x_{N+1}$ until $x_{n+1} < x_b$. The same equilibrium equation is used but the operating line equation is for the stripping section:

$$y_{n+1} = f(x_n) = \frac{\bar{L}}{V} x_n - \frac{Bx_B}{V} \quad (37)$$

Since the last calculated x_{n+1} is not equal to x_b but smaller, is the last ideal stage incomplete. The size of it can be roughly estimated with equation (34).

Calculate the amount of ideal stages in the enriching and stripping sections. The feed plate belongs to stripping section. Notice that during the algorithm the feed plate was calculated to enriching section. Assume that reboiler is one ideal stage and subtract one plate from the amount of the ideal plates in the stripping section. Mark the amounts with $N_{N,E}$ and $N_{N,S}$.

Compare the numbers of ideal stages $N_{G,E}$, $N_{N,E}$, $N_{G,S}$ and $N_{N,S}$ with each other and with the number of actual stages. Calculate the overall tray efficiency

$$\varepsilon = \frac{N}{N_{ACT}} \quad (38)$$

Incorrect estimate as in general instructions, calculate the errors of the material balances numerically.

6. REFERENCES

- McCabe, W.L., Smith, J.C. and Harriot, P., Unit Operations of Chemical Engineering, McGraw-Hill, 5th ed., 1993.
- Oldershaw, C.F., Perforated Plate Columns for Analytical Batch Distillations, Ind. Eng. Chem. Vol 13, No 4, 1991, 265-268

7. NOMENCLATURE

B	bottom product flow, mol/s
D	distillate flow, mol/s
F	feed flow, mol/s
c_p	molar heat capacity, kJ/kmolK
f	function of the operating line
g	function of the equilibrium, $y^*=g(x)$ and $x^*=g^{-1}(y)$
L	total molar flow in liquid, mol/s
i	molar enthalpy of liquid, kJ/kmol
I	molar enthalpy of gas, kJ/kmol
N	number of ideal stages
r	molar latent heat, kJ/kmol
R_B	reboil ratio, dimensionless
R_D	reflux ratio, dimensionless
q	state of the feed, dimensionless
T	temperature, K or °C
V	total molar flow in gas, mol/s
x	mole fraction in liquid, usually for component A, dimensionless
x_C	intersection of operating lines, dimensionless
y	mole fraction in gas, usually for component A, dimensionless
y_C	intersection of operating lines, dimensionless
z	mass transfer ratio for component A, dimensionless

Greek Letters

α	relative volatility, dimensionless
Φ	heat flow, kJ/s

Subindexes

B	bottom product
D	distillate
F	feed
x	liquid
y	gas

Superscripts

x^*	equilibrium value for x
$\bar{L}$	value of L in the bottom of the column

8. APPENDIXES

1. Physical properties of cyclohexane and n-hexane.
2. Refraction indexes for a mixture of cyclohexane and n-hexane.
3. Matching the balances of binary distillation with MS_EXCEL program.

PHYSICAL PROPERTIES OF CYCLOHEXANE AND N-HEXANE

1. Properties of pure substances

Equations for properties:

Density ρ

$$\ln(\rho/1000) = \ln A - (1 - T_r)^{2/7} \cdot \ln B, \text{ where } T_r = T / T_c$$

Heat capacity c_p :

$$c_p = A + BT + CT^2 + DT^3$$

Latent heat ΔH_v :

$$\Delta H_v = \Delta H_{v,ref} \left[(T_c - T) / (T_c - T_{ref}) \right]^n$$

Table 1. Parameters for calculating physical properties of pure substances.

	n-hexane	cyclohexane
Molar weight M	86.178 kg/kmol	84.162 kg/kmol
Boiling point T_{kp}	68.74 °C	80.75 °C
Critical temperature T_c	234.45 °C	281 °C
Latent heat, $\Delta H_{v,ref}$	28.66 kJ/mol	30.33 kJ/mol
in boiling point (T_{ref})		
exponent n	0.381	0.405
Heat capacity c_p	$A = +2.845$	$A = -5.372$
	$B = -11.9 \cdot 10^{-3}$	$B = +56.024 \cdot 10^{-3}$
	$C = +48.703 \cdot 10^{-6}$	$C = -146.858 \cdot 10^{-6}$
	$D = -52.184 \cdot 10^{-9}$	$D = +135.018 \cdot 10^{-9}$
Density	$A = 0.2368$	$A = 0.2729$
	$B = 0.2693$	$B = 0.2727$

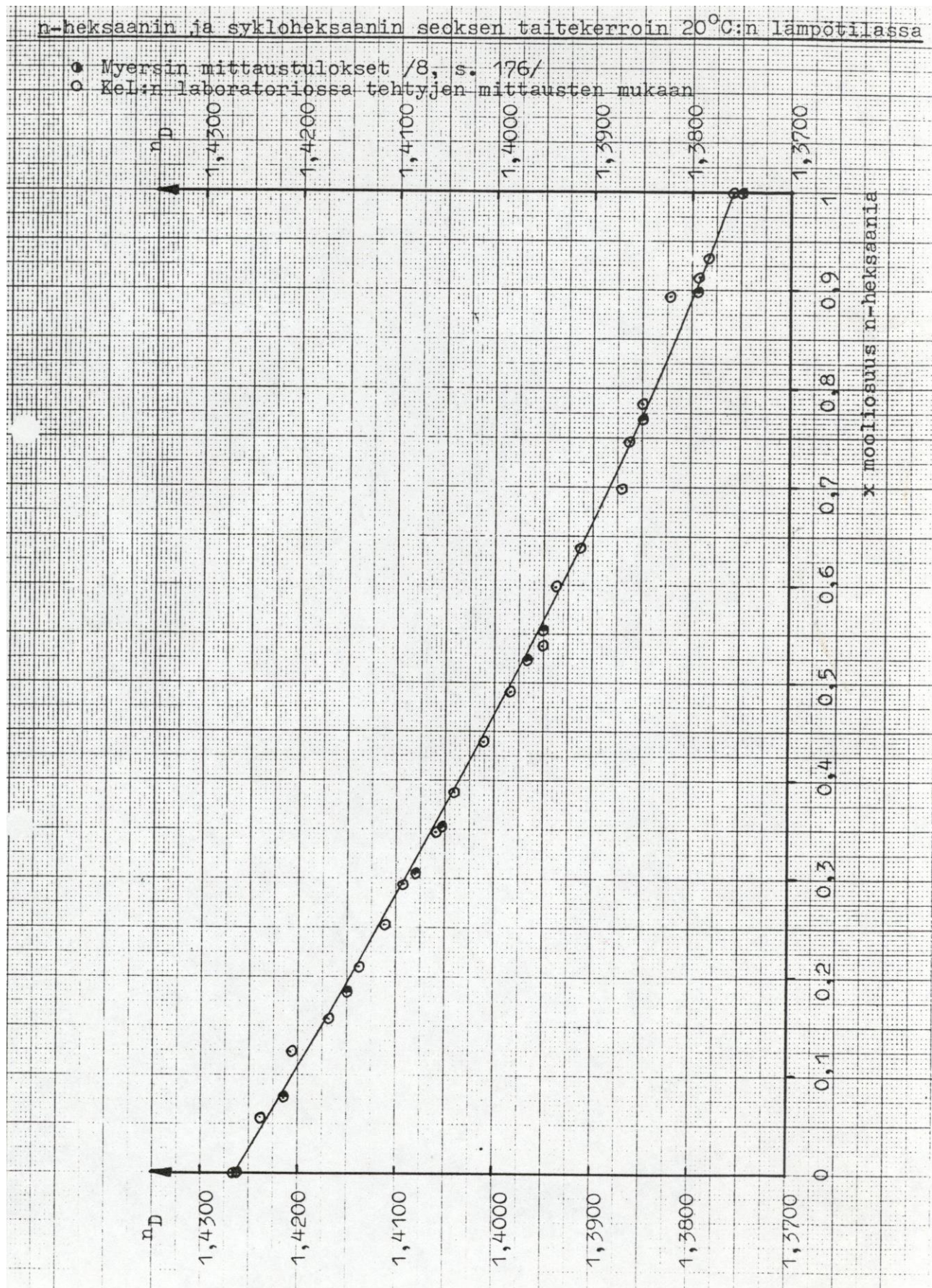
2. Properties for mixture

2.1 Equilibrium

Relative volatility α_{AB} is in the range (1.55, 1.33) when x_A is in the range (0.1, 0.9).

Here, a constant relative volatility is used and $\alpha_{AB}=1.4$

2.2 The refraction index of the mixture in temperature 20 °C is shown graphically in appendix 2.



MATCHING THE BALANCES			8.4.2005
This workbook contains the following sheets:			Color codes
MENU	This sheet.		GIVE
BASIC CASE	Protected therefore Solver does not work.		MATCHED
EXTRACTION	Protected therefore Solver does not work.		CONSTRAINTS
BINARY DISTILLATION	Protected therefore Solver does not work.		TARGET FUNCTION
MACRO	Contains the macros.		
GENERAL PRINCIPLE OF MATCHING THE BALANCES			
The sum $S = \text{SUM}(P_i * ((x_i - X_i) / x_i)^2)$ is minimized			
x_i = measured value of quantity i			
X_i = matched value of quantity i			
P_i = weighting coefficient of quantity i			
so that total and component balances hold true and the sum of mole fractions is one.			
The task at hand is a constrained optimization task, which constraints are the balances and the sum of mole fractions and the minimized function is the weighted sum of relative differences of matched and measured values.			
There is no universal algorithm to solve a problem like this, which would always lead to a solution.			
Therefore matching should be done with few different initial values.			
INSTRUCTIONS			
0 Go to sheet and use the macro CopyPage to make an unprotected sheet.			
1 Give the measured values to cells: MEASURED			
2 If needed give the weighting coefficients to cells: WEIGHT			
The greater the weighting coefficient the smaller is the difference between matched and measured values.			
Weighting coefficient of an accurate measurement should be large.			
Weighting coefficient of an inaccurate measurement should be small.			
Weighting coefficient can be in the range (0, positive infinity). Here the coefficients should be ≥ 1 .			
If there is no special need for weighting, use 1.			
3 Give initial values (e.g. measured values) to cells: INITIAL VALUES			
Leave all other cells UNTOUCHED!			
4 Choose Tools / Solver / Solve: Solver starts the optimization.			
The target cell of the optimization becomes active.			
The constraints are:			
Balances hold true, in other words the cells IN-OUT = 0			
Sum of mole fractions equals to 1, cells SUMx = 1			
All matched values are positive, MATCHED ≥ 0			
5 If the result is not reasonable, try other initial values or change the weighting coefficients.			
Notice that the best result is obtained only by trial and error.			
6 If a reasonable result is not found, save the result and restart solver.			
If a reasonable result is not found after a few times, return to 5 and try other initial values.			
7 Extraction:			
Substance A (component to be extracted) = HAc			
Substance B (diluter) = MIK			
Substance S (solvent) = vesi			
if HAc, of which the feed is made, contains water p -weight %, feed contains $x_B = 1.0 - (1.0 + p/(100-p))x_A$,			
since x_S/x_A is always $p/(100-p)$ in feed, which is made by adding substance B			
In the solvent $y_S = 1.0$ because it is pure water.			
Raffinate and extract are saturated, so x_S is obtained from the saturation curve.			
In raffinate $x_S = 0.56x_A + 0.01$ and in extract $x_S = -1.17x_A + 0.98$.			
In both $x_B = 1.0 - x_A - x_S$			
Extra constraint: in raffinate and extract $x_A \geq 0.001$			
8 Binary distillation			
Notice that $x_B = 1.0 - x_A$			

