

Seminars

Unit Operations II

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I Enthalpy Balances

I.1 Content of methanol in water is to be reduced by a flash distillation unit. The initial methanol content is 30 mol.%; the liquid product contains 20 mol.% of methanol. The solution is fed into the unit at 30 °C and its flow rate is 1 kmol/hr. The pressure is 101.3 kPa. Determine the molar flow rates and composition of both products (waste and distillate) and energy consumption of the unit.

Result: There will be 0.265 kmol/hr of distillate and 0.735 kmol/hr of waste. The distillate contains 57.7 mol.% of methanol. The heat necessary for the distillation is 19.7 MJ/hr.

I.2 2500 kg of a mixture containing 60 wt.% of methanol in water is being separated in a distillation column. From the column, distillate containing 96 wt.% of methanol and bottoms containing 98 wt.% of water are obtained. The reflux ratio is 1.5 and the reflux is returned into the column at its boiling point. The feed is pre-heated from 20°C to its boiling point using a steam of pressure 3 bar, which is also used in the reboiler. Both products are cooled to 25°C using cooling water. The allowed temperature rise of cooling water is 20°C in any of the heat exchangers.

Calculate:

- mass of the products
- consumption of steam and cooling water in the process.

Result: There will be 1543 kg of distillate and 957 kg of bottom product. For the separation is necessary: 2.83 tons of steam and 72.8 tons of cooling water.

I.3 Mixture of ethylene and steam is being cooled in a counter-current heat exchanger from 200°C down to 25°C by cooling water. During the cooling, the steam is condensing and the gas composition changes. At the outlet, the gas is saturated with water vapour and it can be assumed ethylene is insoluble in the condensate. Mass flow rate of the gas is 1200 kg.h⁻¹, 200 kg.h⁻¹ out of which is water. The pressure of the gas is 100 kPa. The inlet temperature of cooling water is 15°C, the outlet temperature is 80°C. Heat losses into surrounding can be neglected. Determine the cooling water consumption.

Result: The cooling water consumption is 3,0 tons per hour.

II Enthalpy Balances with chemical reaction

II.1 One of the routes of nitric oxide (NO) production is partial oxidation of ammonia by air. The reactor is fed with gaseous ammonia (at 25 °C) and air pre-heated to 750 °C. Ammonia conversion is 90%. Maximum temperature at the reactor outlet is 920 °C. The molar ratio of oxygen to ammonia is 2.4 : 1. Determine the amount of heat, which is to be removed per 1 mol of ammonia.

Result: 104 kJ is to be removed per each 1 mol of ammonia.

II.2 Methanol, at 675 °C and 100 kPa, is fed into adiabatic reactor, where formaldehyde is formed by dehydrogenation. Methanol conversion is 25%. Determine outlet temperature of the gas. Assume the heat capacities can be taken constant in the temperature range in question:

$$\begin{aligned}c_P(\text{CH}_3\text{OH}) &= 71 \text{ J}\cdot\text{mol}^{-1}\cdot\text{K}^{-1} \\c_P(\text{HCHO}) &= 50 \text{ J}\cdot\text{mol}^{-1}\cdot\text{K}^{-1} \\c_P(\text{H}_2) &= 29 \text{ J}\cdot\text{mol}^{-1}\cdot\text{K}^{-1}\end{aligned}$$

Result: The outlet temperature is 365 °C.

II.3 By complete burning of 1 kg of coal, 8 kg of steam at pressure 1.2 MPa is produced. Temperature of water entering the steam generator is 40 °C.

- What amount of steam at pressure 1,6 MPa can be produced using 1 kg of coal, when inlet water temperature is 10 °C.
- Calculate the amount of steam at pressure 1.2 MPa, which would be produced by burning 1 kg of coal in 50% excess of air. The temperature of inlets is 25°C and temperature of exhaust gasses is 500°C. Compare the result with the number given in the problem.

Result: At these conditions, 7.6 kg of steam at 1,6 MPa will be produced.

II.4 What amount of dry saturated steam of temperature 120°C could be theoretically obtained by burning 1 tonn of sulphur using stoichiometric amount of air, assuming the product is sulphur dioxide. Temperature of water and sulphur vapour at the inlet is 25°C, temperature of the outlet gas is 400°C.

Result: It is possible to obtain 2.85 tonns of steam at conditions required.

Additional values:

Heat of vaporization

$$\Delta h_{\text{vap}}(\text{methanol}, 81.8^\circ\text{C}) = 34374.4 \text{ J / mol}$$

$$\Delta h_{\text{vap}}(\text{water}, 81.8^\circ\text{C}) = 41519.5 \text{ J / mol}$$

Heat capacity of liquid

$$c_p(\text{methanol}) = 94.04 - 0.2352 T + 5.0109 \cdot 10^{-4} T^2 + 4.7421 \cdot 10^{-7} T^3 \text{ J/mol.K } \{T \text{ in K}\}$$

$$c_p(\text{water}) = 232.20 \text{ J/mol.K}$$

III Sedimentation

III.1 Spherical particle, density 3320 kg.m^{-3} , settles in a 30 wt.% solution of methanol in water at 40°C . It was measured, that it passes 1,6m within 52s. Determine the terminal settling velocity in air at 90°C and normal pressure. What is the diameter of the particle?

Result: The particle diameter is 0.2 mm. The terminal settling velocity in air is 1.71 m.s^{-1} .

III.2 Determine minimal diameter of particles with density 1600 kg.m^{-3} , which will not be entrained by air flow from a fluid dryer. Air at normal pressure and 80°C is used for drying. The cross-section of the dryer is 0.5 m^2 and volumetric gas flow rate is $12000 \text{ m}^3/\text{hr}$.

Result: The minimum diameter of the particles is 1.5 mm.

III.3 Determine the terminal sedimentation velocity of a spherical particle with diameter 0.1 mm and density 2400 kg.m^{-3} in a gas of density 1.25 kg.m^{-3} and viscosity $16 \text{ }\mu\text{Pa.s}$.

Result: The terminal sedimentation velocity is 0.592 m.s^{-1} .

III.4 Determine the terminal sedimentation velocity of a spherical particle with diameter 0.1 mm and density 2400 kg.m^{-3} in a gas of density 1.25 kg.m^{-3} and viscosity $16 \text{ }\mu\text{Pa.s}$.

Result: The terminal sedimentation velocity is 0.592 m.s^{-1} .

III.5 What is the necessary diameter of a vertical cylindrical thickener for separation of solids from $50 \text{ m}^3/\text{hr}$ of water suspension of temperature 20°C . The suspension contains 10 wt.% of particles with density 3200 kg.m^{-3} and diameter $10 \text{ }\mu\text{m}$. Solids are concentrated into a suspension containing 50 wt.% of solids.

Result: The settler diameter is 11.3 m.

III.6 Determine the least diameter of particles, which will be separated from gas in a dust chamber. The dust chamber is 6 m long, 2m wide and contains 45 plates spaced 60 mm apart. The flow rate of gas with particles is 5000 kg.s^{-1} , its density in the dust chamber is 0.51 kg.m^{-3} and its dynamic viscosity is $32.75 \text{ }\mu\text{Pa.s}$. Determine also, how the gas flow rate could be increased, if the dust particles were larger than $15 \text{ }\mu\text{m}$.

Result: Particles larger than $7.78 \text{ }\mu\text{m}$ will be separated. For particles larger than $15 \text{ }\mu\text{m}$, the gas flow rate would increase to 13550 kg.h^{-1} .

III.7 A dust chamber removes all particles with diameter $90\ \mu\text{m}$ from $2500\ \text{m}^3\cdot\text{hr}^{-1}$ of air at 40°C . The density of the particles is $2100\ \text{kg}\cdot\text{m}^{-3}$. This chamber is to be used for sedimentation of particles with density $2910\ \text{kg}\cdot\text{m}^{-3}$ and diameter $30\ \mu\text{m}$ from air at 250°C .

Calculate:

- the flow rate of air, which allows sedimentation of all particles
- least diameter of particles, which will be quantitatively removed at air flow rate $2500\ \text{m}^3\cdot\text{hr}^{-1}$.

Result: The flow rate of air is $265\ \text{m}^3\cdot\text{hr}^{-1}$. At $2500\ \text{m}^3\cdot\text{hr}^{-1}$, particles with diameter higher than $75\ \mu\text{m}$ will be removed.

Definition of criteria used in sedimentation:

$$Re_u = \frac{u_p d_p}{\nu_l} = \frac{u_p d_p \rho_l}{\eta_l}$$

$$Ar = \frac{f d_p^3 |\rho_p - \rho_l|}{\rho_l \nu_l^2} = \frac{f d_p^3 \rho_l |\rho_p - \rho_l|}{\eta_l^2}$$

$$Ly = \frac{Re_u^3}{Ar} = \frac{u_p^3 \rho_l^2}{f |\rho_p - \rho_l| \eta_l}$$

Equations for sedimentation of spherical particles		
Region	Equation	Validity region
Laminar (Stokes)	$\zeta_u = 24 / Re_u$ $Re_u = Ar/18$ $Re_u = \sqrt{(18 Ly)}$ $Ly = Ar^2 / 5832$	$Re_u \leq 0.2$ $Ar \leq 3.6$ $Ly \leq 2.22 \times 10^{-3}$ $Ar \leq 3.6$
Laminar and transient (Allen)	$\zeta_u = 24(1 + 0.125 Re_u) / Re_u^{0.72}$ $Ar = 18 Re_u (1 + 0.125 Re_u)^{0.72}$ $Ly = Re_u^2 / (18(1 + 0.125 Re_u^{0.72}))$ $Ar^{2/3} = 18 Ly^{1/3} [1 + 0.125 (Ar Ly)^{0.24}]$	$Re_u \leq 10^3$ $Ar \leq 3.43 \times 10^5$ $Ly \leq 2.92 \times 10^3$ $Ar \leq 3.43 \times 10^5$
turbulent (Newton)	$\zeta_u = 0.44$ $Re_u = 1.73 \sqrt{Ar}$ $Re_u = 0.33 Ly$ $Ar = 3.6 \times 10^{-2} Ly^2$	$Re_u \in \langle 1000; 1.5 \times 10^5 \rangle$ $Ar \in \langle 3.43 \times 10^5; 7.4 \times 10^9 \rangle$ $Ly \in \langle 2.92 \times 10^3; 4.6 \times 10^5 \rangle$ $Ly \in \langle 2.92 \times 10^3; 4.6 \times 10^5 \rangle$
transient (lower precision)	$\zeta_u = 18.5 Re_u^{-0.6}$ $Re_u = (Ar / 13.9)^{0.714}$ $Re_u = (13.9 Ly)^{0.625}$ $Ar = 138.6 Ly^{0.875}$	$Re_u \in \langle 0.4; 500 \rangle$ $Ar \in \langle 3.85; 8.33 \times 10^4 \rangle$ $Ly \in \langle 1.66 \times 10^{-2}; 1500 \rangle$ $Ly \in \langle 1.66 \times 10^{-2}; 1500 \rangle$

IV Residence time distribution

IV.1 Determine the mean residence time distribution in apparatus. Measuring a response characteristics on pulse injected at the inlet, following concentrations at the outlet were measured:

τ (min)	0.1	0.2	1.0	2.0	5.0	10.0	30.0	>30
c (kg/m ³)	0.60	0.51	0.45	0.375	0.21	0.06	0.003	0

Show, which ideal type of flow is closer to the flow in the apparatus.

Result: The mean residence time in the apparatus is 4.75 minutes.

IV.2 Determine the mean residence time in the apparatus, where following concentrations were measured after a step-change of concentration at the inlet:

τ (s)	0	15	25	35	45	55	65	75	95	115
c (kg/m ³)	0	0.5	1.0	2.0	4.0	5.5	6.5	7.0	7.7	7.9

Create a plot of the residence time distribution curve at the system outlet $F(\tau)$ and determine mean residence time.

Result: The mean residence time in the apparatus is approximately 47 s.

IV.3 Table below provides waiting time of drivers by 23 petrol stations. What is the mean waiting time?

total waiting time (min)	0	3	6	9	12	15	16	21
Number of petrol stations with given waiting time	0	4	3	5	8	2	1	0

Result: The mean waiting time is 9.5 minutes.

IV.4 180 m³/hr of gas flows through a packed column with a cross-section of 1.3 m². The packing is made of ceramic Rashig rings 15mm in diameter. The design height of the packing is 5m. What would be the height of the packing, if the influence of axial dispersion was taken into account assuming the process requires 10 transfer units for the separation. Liquid flow rate effects can be neglected.

Result: The corrected height of the packing will be 7.14m.

V Reactors

Batch reactors

V.1 Butyl acetate is produced in a batch reactor from acetic acid and butanol. Determine the reaction time for 50 % conversion of the acid. Initially, the batch contains butanol and acetic acid (compound A) in molar ratio 4.97 : 1. Calculate the volume of the reactor and amount of each component which is necessary for production of 50 kg.h⁻¹ of ester. Time for cleaning and refilling is 45 min. The reaction rate is described by $r = k c_A^2$, where the reaction constant $k = 1.74 \cdot 10^{-2} \text{ m}^3 \cdot \text{kmol}^{-1} \cdot \text{min}^{-1}$. The density of the reaction mixture can be taken as constant during the reaction period. Density of acetic acid and butanol is 958 kg.m⁻³ and 742 kg.m⁻³, respectively.

Assuming constant density, the initial concentration of acetic acid can be calculated as follows:

$$c_{A0} = \frac{n_{A0}}{V_{A0} + V_{B0}} = \frac{n_{A0}}{m_{A0} / \rho_A + m_{B0} / \rho_B} = \frac{n_{A0}}{n_{A0} M_A / \rho_A + n_{B0} M_B / \rho_B}$$

Result: The volume of the reactor is 0.62 m³. Feed contains 66.6 kg of acetic acid and 409 kg of butanol. The reaction time is 32.1 min.

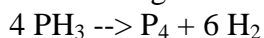
V.2 Liquid phase reaction $A + B = C$ is carried in a batch reactor. The feed temperature is 293 K. Following data are available:

$$c_{A0} = 5 \text{ kmol} \cdot \text{m}^{-3}, \Delta h_r = -21.96 \text{ kJ} \cdot \text{mol}^{-1}, \rho = 830 \text{ kg} \cdot \text{m}^{-3}, c_p = 1.257 \text{ kJ} \cdot \text{kg}^{-1} \cdot \text{K}^{-1}$$

The density and heat capacity of the mixture are constant. In a range $T = 293 \div 393 \text{ K}$, determine and plot dependence of the conversion of species A on temperature at which the reactor is operated. Assume adiabatic operation of the reactor.

Result: At 393 K, the conversion of A is 95%.

V.3 500 g of gaseous phosphine is decomposed in a batch reactor at constant volume at 945 K and 101 kPa during 200s.



The reaction is endothermic, irreversible and of first order. The rate constant is a function of temperature:

$$\log k = -\frac{18963}{T} + 2 \log T + 11.53$$

Where T is temperature in K.

Determine: a) mass of phosphorus obtained (in g)
b) time to get 95% conversion of at given temperature.

Result: a) 399 g of phosphorus is formed in 200s; b) 288 s are needed to get 95% conversion.

V.4 A batch reactor is used for reaction $A + B \rightarrow C$. The rate of the reaction is given $r = k c_A c_B$. Calculate reactor volume for processing 4400 tons of reaction mixture of A + B per year, while reaching 90% conversion of A. The density of A is 830 kg.m^{-3} , its molar weight 46 kg.kmol^{-1} ; density of B is 800 kg.m^{-3} , its molar weight 58 kg.kmol^{-1} ; density of C is 950 kg.m^{-3} , its molar weight 104 kg.kmol^{-1} . The weight ratio m_{B0}/m_{A0} in the reaction mixture is 2.5; $k = 8.33 \cdot 10^{-5} \text{ m}^3\text{kmol}^{-1}\text{s}^{-1}$. Time for emptying and refilling of the reactor 1800s.

Determine: The volume is $1,12 \text{ m}^3$.

Result: a) 399 g of phosphorus is formed in 200s; b) 288 s are needed to get 95% conversion.

Continuously stirred tank reactors (CSTR)

V.5 Reaction $2A \rightleftharpoons B + C$ is carried out isothermally in a continuous stirred tank reactor (CSTR). Volumetric flow of the reaction mixture is $V = 100 \text{ m}^3.\text{h}^{-1}$. Concentration of component A at the input to the cascade is $c_{Ai} = 1.5 \text{ kmol.m}^{-3}$, input concentration of components B and C is $c_{Bi} = c_{Ci} = 0$. The rate constant of the forward reaction is $k = 5 \text{ m}^3.\text{kmol}^{-1}.\text{h}^{-1}$, the equilibrium constant is $K = 16.0$. Conversion at the output from the reactor is 80 % of the equilibrium conversion. The kinetic equation for the reaction is give as $r = k (c_A^2 - c_B c_C / K)$.

Calculate the volume of the reactor, which is needed to achieve conversion of $0.8\zeta_{\text{eq}}$.

Determine the number of reactors of 1/10 volume of the one from previous case needed for the same task.

Result: The volume of a single reactor is 62.8 m^3 . 4 reactors are needed.

V.6 500 tons of styrene per year¹ is used in production of butadien-styren copolymer. A CSTR (continuous stirred tank reactor), the volume of which is 1.8 m^3 , is available. The feed contains 92 mol.% of styrene and its density is 870 kg.m^{-3} . The conversion of styrene should be 30%. The reaction rate of styrene can be expressed as $r_S = -k.c_S^{3/2}$, where the reaction kinetic constant is a function of temperature $k = A \exp(-E_a/RT)$. The activation energy E_a for the reaction is $96.79 \text{ kJ.mol}^{-1}$ and the pre-exponential factor $A = 3.3 \cdot 10^{11} \text{ m}^{1.5}\text{kmol}^{-0.5}\text{h}^{-1}$.

Determine the temperature, at which the reactor should operate.

Result: The required temperature is 370K (97°C).

¹ For maintenance reasons, during 1 year only 8000 h are available for uninterrupted operation of the production units.

V.7 Acetic anhydride is to be hydrolysed in a series of 4 CSTR's (continuous stirred tank reactor). The reactors are having identical volume, but will be operated at different temperature, according to the following table:

Reactor No.	1.	2.	3.	4.
Temperature [K]	283	288	298	313
Rate constant [10^{-3} s^{-1}]	0.96	1.34	2.63	5.22

The volumetric flow rate of the feed mixture is 1.67 l.s^{-1} and it contains 0.9 mol.l^{-1} of acetic anhydride. The reaction rate is given as $r = kc_A$, where A is the acetic anhydride.

Determine volume of the reactors if overall conversion of A should be 91%.

Determine number of these reactors in the series, if all were operating at temperature of 288K.

Result: The volume of the reactors is 0.6 m^3 . 6 reactors operating at 288K would be required.

Plug flow reactors (PFR)

V.8 A reaction in a liquid phase with constant density $2A + B \rightarrow 2C$ proceeds according to the expression $r = kc_A^2$ in an isothermal tubular reactor (with plug flow). Initial concentrations are: $c_{A0} = c_{B0} = 1.5 \text{ kmol.m}^{-3}$, $c_{C0} = 0 \text{ kmol.m}^{-3}$. 95 % conversion of A is required. The feed flow rate is $0.1 \text{ m}^3.\text{h}^{-1}$ and the rate constant is $0.01 \text{ m}^3.\text{h}^{-1}.\text{mol}^{-1}$. The internal cross-sectional area is 0.002 m^2 .

a) What is the flow of component C at the output from the reactor?

b) What is the length of the reactor? The internal cross-sectional area is 0.002 m^2 .

Result: a) The flow of component C at the output is $0.143 \text{ kmol.h}^{-1}$
b) The length of the reactor is 31.7 m.

V.9

Sulphuryl chloride decomposes according to the scheme $\text{SO}_2\text{Cl}_2 \rightarrow \text{SO}_2 + \text{Cl}_2$. At a temperature of 320°C the reaction is irreversible and has a rate constant $2.2 \cdot 10^{-3} \text{ s}^{-1}$. Calculate the volume of a tubular flow reactor needed for 90% decomposition. The pressure is 101 kPa, the feed of SO_2Cl_2 is 50 g h^{-1} .

Result: The reactor volume is 8.43 m^3 .

V.10 Sucrose ($\text{C}_{12}\text{H}_{22}\text{O}_{11}$) is enzymatically hydrolyzed to glucose and fructose in a pipe reactor. The reactor capacity should be 5 t of sucrose per day. Concentration of sucrose in the inlet solution is 1 kmol.m^{-3} . The conversion should be at least 95%. The reaction rate is given in a form:

$$r = \frac{k_1 w_e c_s}{1 + k_2 c_s + k_3 c_s^2}$$

where the rate constants k_1 through k_3 have at the reaction conditions following values: $k_1 = 5 \times 10^{-3} \text{ m}^3 \text{ kg}^{-1} \text{ s}^{-1}$, $k_2 = 3 \times 10^{-3} \text{ m}^3 \text{ mol}^{-1}$, and $k_3 = 5 \times 10^{-4} \text{ m}^6 \text{ mol}^{-2}$. Concentration of biocatalyst w_e is $50 \text{ kg} \cdot \text{m}^{-3}$. The reaction rate includes so called substrate inhibition – reduction of hydrolysis rate at high substrate concentration. Assume the density of the solution does not change.

Determine:

- a) Volume of a pipe reactor.
- b) Volume of the reactor at sucrose inlet concentration twice as high.
- c) Volume of an CSTR for hydrolysis of solution with sucrose concentration 2 kmol m^{-3} .

Results:

- a) $0,1727 \text{ m}^3$
- b) $0,3405 \text{ m}^3$
- c) $0,0405 \text{ m}^3$

VI Liquid – liquid extraction

VI.1 Acetic acid is to be extracted from 30 kg of a mixture containing 75 wt.% of water and 25 wt.% of acetic acid using diethyl ether. The concentration of acetic acid in raffinate has to be below 11.5 wt.%. In each stage, 7kg of diethyl-diethyl ether is added. Determine the required number of stages and the required amount of extraction solvent, if equilibrium is reached in each stage. The extraction is performed at 25°C.

Result: The required number of stages is 4 and 28kg of solvent is required

VI.2 A solution of 30 wt.% acetic acid in water is extracted counter currently with pure diethyl ether such that the concentration of acetic acid in water drops to 3 wt.% and the extract contains 17 wt.% acetic acid. Determine the required height of the packing and the required amount of diethyl ether per 1 kg of feed. The extraction is performed at 25°C. The height equivalent to a equilibrium plate is 0.75m.

Result: The required height of packing is 2.8 m and 1 kg of feed requires 1.26 kg of solvent.

VI.3 a) What will be the mass and concentration of phases after mixing and attainment of equilibrium, if we mix 5 kg of a mixture containing 90 wt.% aniline and 10 wt.% cyclohexane with

1. 2 kg cyclohexane
2. 2 kg heptane

b) How much aniline, contaminated with 8 wt.% of cyclohexane and 2 wt.% heptane, must be added to 1 kg of solution containing 50 wt.% heptane and 50 wt.% cyclohexane, such that 2 phases are obtained? What will be the compositions of these two phases? The process occurs at 25°C.

Results:

Part of question	Phase	Mass (kg)	Constituents (wt.%)		
			cyclohexane	Aniline	heptane
a1	Hydrocarbon	1.16	83.5	16.5	0
	Aniline	5.84	26.2	73.8	0
a2	Hydrocarbon	2.34	18.4	8.1	73.5
	Aniline	4.66	1.5	92.5	6.0
b	Hydrocarbon		45.2	10.4	44.4
	Aniline		7.2	88.0	4.0

- VI.4 Acetone is to be extracted from a 5 kg mixture of 60 wt.% acetone and 40 wt.% water by a trichloroethylene solvent containing 16 wt.% of acetone.
- Determine the composition and mass of the extract and raffinate, if following amounts of solvent is added: 0.5 kg, 5 kg, 50 kg.
 - Determine the minimum mass of the solvent, which is required for the formation of two phases.
 - Determine the maximum mass of solvent, which is required for the existence of two phases.

Assume equilibrium is reached.

Result: a) 1. 5.5 kg of a single phase of concentration 56 wt.% acetone, 7.6 wt.% trichloroethylene, and 36.4 wt.% water a) 2. and a) 3. in the table below

b), c) The minimum mass of extraction agent is 0.7 kg, the maximum is 274 kg

Part of question	stream	Mass (kg)	Constituents (wt.%)		
			acetone	trichloroethylene	water
a2	raffinate	2.61	29.1	1.1	69.8
	extract	7.39	40.9	56.5	2.6
a3	raffinate	1.78	13.4	0.7	85.9
	extract	53.22	20.2	78.9	0.9

- VI.5 A 20 mol.s⁻¹ mixture containing 55 mol.% trichloroethylene, 42 mol.% methanol and 3 mol.% mol water is extracted counter currently by a mixture containing 12% mol methanol and 88 mol.% water. The extraction is performed at 20°C. What is the required amount of extraction agent and the number of equilibrium stages, if the required concentration of methanol in the raffinate is 1 mol.% and in the extract 39 mol.%.

Result: 17.5 mols⁻¹ of extraction agent and 2 equilibrium stages are required for the extraction.

VII Absorption in stage columns

VII.1 Benzene is stripped from a coke oven gas in an absorption column. Volumetric flow rate of the gas is $850 \text{ m}^3\text{h}^{-1}$. From the gas feed, 95% of the benzene is to be removed. The pressure at the inlet point of the column is 1050 Torr, the temperature in the column is 30°C . The vapour pressure of the oil at this temperature is negligible. Washing oil is used as a stripping solvent and its mean molar weight is 260 kg kmol^{-1} . The concentration of benzene in the inlet oil is 0.5% mol and in the inlet gas is 2% mol. For absorption, 1.5 times the minimum required amount of solvent is used. Assume an ideal solution of benzene in oil, the saturated vapour pressure of benzene at 30°C is 15.91 kPa. Calculate the final concentration of benzene in oil.

Result: The outlet oil from the column contains 11.9% mol benzene. The number of theoretical plates is 7.1.

VII.2 10 tonns.hr⁻¹ of air containing SO₂ enters an absorption column. The inlet contains 10 wt.% of SO₂, from which 95 % is to be removed by absorption into pure water. The column has 21 plates and the column efficiency is 30 %. The equilibrium line can be expressed as $Y = 22.1 X$, where Y and X are relative mol. fractions of SO₂ in gas and liquid respectively. Determine the mass flow rate of water and outlet concentration of SO₂ in the water.

Result: The mass flow rate of water is 163 tonns.hr⁻¹. The concentration of SO₂ in water at the exit is 0.583 wt.%.

VII.3 Gas is fed into an absorption column at superficial velocity 640 m.hr^{-1} at 20°C and 3800 Torr. The gas contains 3 vol.% of NH₃, the rest consists of inert gases. The gas is contacted with water, which contains 0.5 wt.% of NH₃ at the inlet. The flux intensity of water is $1200 \text{ kg.m}^{-2}.\text{hr}^{-1}$. 92 % of the ammonia is to be removed from the gas. Determine the number of theoretical stages needed for the task. Determine the number of real stages, if the Murphree efficiency E_y for individual stages is 0.65.

Result: Number of theoretical stages is 2.3, the number of real stages is 4.5.

VII.4 Effluent gas from a reactor contains 3 mol.% of ethylenoxide (compound A), the rest being carbon dioxide and nitrogen. Ethylenoxide is to be absorbed to water in counter-current packed column operated at 2 MPa and 30°C . 95 % of ethyleneoxide is to be absorbed into water. Inlet water contains 0.1 mol.% of ethylenoxide, 45 kmol.h^{-1} of gas is processed and molar ratio of H₂O to inert gases is 1:1. The superficial gas velocity at the bottom of the column is 0.02 m.s^{-1} . Height equivalent to a theoretical plate is 0.5 m. At the conditions, the solubility of CO₂ and N₂ in water and water evaporation may be neglected. Calculate:

- a) height of packing
- b) column diameter
- c) ratio between real consumption of the solvent and minimal consumption.

The equilibrium is characterized by linear equation $Y_A = 0.65 X_A$, where Y_A and X_A are relative mole fractions.

Result: The packing height in the adsorption column is 3 m and diameter is 1 m. The ratio of real to minimum solvent consumption is 1.6.

VII.5 Determine the consumption of diethanolamine solution, which is needed for absorption of CO_2 from air. Inlet gas contains 17.5 vol.% of CO_2 , 95% of which is to be absorbed. The gas flow rate at the inlet is $3000 \text{ kmol}\cdot\text{hr}^{-1}$ and the absorption is carried out at 25°C and 1 bar. The absorption solution contains $2 \text{ mol}\cdot\text{dm}^{-3}$ of diethanolamine and its density is $1024 \text{ kg}\cdot\text{m}^{-3}$. The flow rate of the absorption solution should be by 25% higher than the minimal consumption.

Result: The consumption of the solution is $264 \text{ mol}\cdot\text{s}^{-1}$.

VII.6 Propane is removed from absorption oil by desorption in a plate column. The oil contains 11 mol.% of propane at the inlet and 0.5 mol.% at the outlet. How many theoretical stages are needed for the desorption, if the steam consumption is 1.5 times the minimum consumption? The equilibrium relationship for the conditions in the column is given as $y = 0.526 x$ (where y and x are mol fractions of propane).

Result: The column should have 5.7 theoretical stages.

VII.7 Air containing acetone vapour is fed into an absorption column. The content of acetone at the inlet is 2.1 kg of acetone per 1 kg of air. 95% of the acetone entering the column is to be removed by absorption into pure water. The interfacial velocity at the column inlet is $0.5 \text{ m}\cdot\text{s}^{-1}$. The slope of the operating line at minimum water consumption conditions is equal to 1.4 (for the case of concentration given in relative weight fractions). The pressure in the column is 2100 Torr. Relative weight fraction of acetone in the water at the exit is 50% of the theoretical maximum. 200 kg of acetone vapour per hour enters the column. The temperature inside the column is 20°C . Determine the flow rate of water.

Result: The flow rate of water is $267 \text{ kg}\cdot\text{h}^{-1}$.

Material balance of absorption column

Material balance of an absorption column can be given as:

$$y_{AGi}n_{Gi} + x_{ALi}n_{Li} = y_{AGe}n_{Ge} + x_{ALe}n_{Le}$$

Because the flow rates of gas and liquid are changing through the column solely due to absorption of the solute A, the flow rates of the other species of the gas and liquid are constant:

$n_B = (1-y_{AG})n_G$; $n_C = (1-x_{AL})n_L$. By substituting this into the material balance:

$$\frac{y_{AGi}}{1-y_{AGi}}n_B + \frac{x_{ALi}}{1-x_{ALi}}n_C = \frac{y_{AGe}}{1-y_{AGe}}n_B + \frac{x_{ALe}}{1-x_{ALe}}n_C$$

Where the fractions can be taken as 'relative fractions':

$$Y_i n_B + X_i n_C = Y_e n_B + X_e n_C$$

From this relationship, an equation of an operating line can be derived:

$$Y_{k+1} = \frac{n_C}{n_B}(X_k - X_i) + Y_e$$

Here, the symbol k denotes an arbitrary theoretical stage in the column, starting the indexing from the top.

Numerical solution of absorption column for linear equilibrium

If the equilibrium relationship can be approximated using a linear relationship for the whole column, it is possible to modify the material balance equation using absorption factor:

$$\zeta_Y = \frac{n_C}{\varphi_A n_B} \quad \text{or} \quad \zeta_X = \frac{\varphi_A n_B}{n_C} = \frac{1}{\zeta_Y}$$

And absorption factor:

$$\eta_Y = \frac{Y_{Ai} - Y_{Ae}}{Y_{Ai} - Y_{AE}^*} \quad \text{or} \quad \eta_X = \frac{X_{Ae} - X_{Ai}}{X_{Ae}^* - X_{Ai}} = \frac{\eta_Y}{\zeta_Y}$$

For counter current absorber with *NTP* theoretical stages, the material balance takes form

$$\zeta_Z^{NTP} = \frac{1 - \eta_Z}{1 - \eta_Z} \quad \text{where } Z = X, Y$$

For the case of $\zeta = 1$, the relationship simplifies to

$$NTP = \frac{\eta_Z}{1 - \eta_Z}$$

VIII Heat transfer

Free convection

VIII.1 Determine the heat losses from a non-isolated horizontal pipeline for supplying steam at 1 MPa. Ambient temperature is 20°C and air pressure is 98 kPa. The pipe is 50 m long and has outer diameter 51mm. Heat losses due to radiation and heat transfer resistance of the pipe wall are negligible.

Result: The heat losses are 12.9 kW.

Phase change

VIII.2 Saturated steam is condensing on a horizontal pipe, which is 1m long and has outer diameter 20mm.

Calculate the heat transfer coefficient and mass flow rate of condensate produced in 1 hour when:

- a] pressure of steam is 5 kPa and wall temperature is 27.1°C,
- b] pressure of steam is 100 kPa and wall temperature is 90.4°C,

Result: a) $\alpha = 1.13 \cdot 10^4 \text{ W}\cdot\text{m}^{-2}\cdot\text{K}^{-1}$; $m = 6.12 \text{ kg}$.
b) $\alpha = 1.34 \cdot 10^4 \text{ W}\cdot\text{m}^{-2}\cdot\text{K}^{-1}$; $m = 12.3 \text{ kg}$.

VIII.3 Calculate the heat transfer coefficient for ethanol vapour condensing at atmospheric pressure on a surface of a horizontal pipe, which is 1.2 m long and 20mm in outer diameter. Experimentally, 0.01 kg·s⁻¹ of condensate was obtained and the mean temperature of cooling water is 61.7°C. The heat resistance of the pipe wall can be neglected.

Result: The heat transfer coefficient is 2.3 kW·m⁻¹K⁻¹.

VIII.4 Water boils in a kettle at 100kPa on a copper tube, 500 mm long and with diameter 30 mm. The heating inside of the tube is producing 2.5 kW of energy. Assuming all the heat flux is transferred into the water, calculate the outer wall temperature.

Result: The outer wall temperature will be approximately 110°C.

Calculation of heat transfer coefficient:

Boiling

One of approximate relationships for heat transfer coefficient calculation is

$$\alpha = Kq^r p^z$$

where q is heat flux [kJ/m²] and p pressure. K , r and z are empirical constants. For clean and smooth surface $r=0.7$; $z=0.4$. Constant K depends on liquid:

water: $K=2.4 \cdot 10^{-2}$	methanol: $K = 0.85 \cdot 10^{-2}$
ethanol: $K=1.1 \cdot 10^{-2}$	benzene: $K = 0.74 \cdot 10^{-2}$

Condensation

For film condensation of saturated vapour and laminar flow of condensate, Nusselt derived following relation:

$$\alpha = C \left[\frac{\lambda^3 \rho^2 g \Delta h_{lv}}{l \eta (t_v - t_w)} \right]^{0.25}$$

where the constant C is 1.15 for condensation on flat vertical surface (and l is then vertical dimension of the wall) or 0.725 for condensation on horizontal pipe (and l is pipe diameter). Properties of the condensate (density, viscosity and heat conductivity) are to be substituted at average temperature of the wall and the vapour.

Free convection

In case, the dimensions of the space around the heat transfer area is much larger than the heat transfer area itself, a simple relationship may be used:

$$Nu = C (GrPr)^n$$

where C and n are empirical constants given in table below

$GrPr$	C	n	error estimaion [%]
$<10^{-2}$	0.5	0	± 15
$10^2 - 5 \cdot 10^2$	1.18	0.125	± 15
$5 \cdot 10^2 - 2 \cdot 10^7$	0.54	0.25	± 15
$2 \cdot 10^7 - 5 \cdot 10^{13}$	0.135	1/3	± 8

The characteristic dimension is

- diameter – for sphere or horizontal cylinder
- height – for vertical flat surface

IX Mass transfer by diffusion and convection

IX.1 Benzene is stored in a storage tank with inner diameter of 2 m. During maintenance check the cover was removed for 36 hours while keeping the liquid inside, the level being 10 cm below the top of the tank. The temperature was 15°C and atmospheric pressure 100 kPa. The saturated vapour pressure of benzene at these conditions is 8 kPa. Assuming the air layer above the level is stagnant (not moving) and benzene concentration above the tank is negligible, determine approximate amount of benzene lost due to evaporation.

Result: The losses of benzene are about 9.08 kg.

IX.2 A mixture of methanol and water is boiling in an experimental apparatus at temperature 80°C and pressure 100 kPa. 30 cm above the liquid surface is the partial pressure of methanol 30 kPa. The surface area is 30 cm². Assuming the vapours above the liquid are stagnant and that the total flux through the liquid surface is negligible; calculate the total flow rate of methanol through the interface.

Result: The total flow rate of methanol through the interface is $2.97 \cdot 10^{-6} \text{ mol.s}^{-1}$.

IX.3 Pharmaceutical product must be protected against oxidation by atmospheric oxygen. Two options of wrapping material have been suggested: either polyethylene foil with thickness 1.6mm or polyamide foil with thickness 0.8mm. Maximum allowable partial pressure of oxygen inside the packaging is 100 Pa, air pressure is 100 kPa and maximum temperature is 30°C. Oxygen permeability in polyethylene is $2.86 \cdot 10^{-16} \text{ m}^2\text{s}^{-1}\text{Pa}^{-1}$, permeability of oxygen in polyamide is $4.12 \cdot 10^{-18} \text{ m}^2\text{s}^{-1}\text{Pa}^{-1}$. Estimate the flux of oxygen through these materials and choose the more suitable one.

Result: The flux of oxygen through polyethylene foil is $1,48 \cdot 10^{-7} \text{ mol.m}^{-2}\text{s}^{-1}$; the flux of oxygen through the polyamide foil is $4.27 \cdot 10^{-9} \text{ mol.m}^{-2}\text{s}^{-1}$. Polyamide foil wrapping will be more suitable.

IX.4 SO₂ is being absorbed from air into a thin film of water in a wetted wall column – a pipe, which is 1.5 m long and 0.1 m in diameter. The mean gas velocity is 2 m.s⁻¹. The temperature is 15°C and the pressure is 98 kPa.

Calculate the mass transfer coefficient in gas phase using relation:

$$\mathbf{Sh} = 0.023 \mathbf{Re}^{0.83} \mathbf{Sc}^{0.44}$$

Where

Sherwood number	$\mathbf{Sh} = k_c d / D_{AB}$
Reynolds number	$\mathbf{Re} = u_G d / \nu$
Schmidt number	$\mathbf{Sc} = \nu / D_{AB}$

Result: The gas-side mass transfer coefficient is $8.15 \cdot 10^{-3} \text{ m.s}^{-1}$.

IX.5 Calculate the overall mass transfer coefficient for absorption of SO_2 into water in the wetted wall column from problem IX.4. The liquid side mass transfer coefficient is $k_x = 8.9 \text{ mol.m}^{-2}\text{s}^{-1}$. The driving force should be expressed in mol fractions in gas phase. The equilibrium relation in mol fractions can be simplified to $y = 7.979 x$.

Result: The overall mass transfer coefficient $K_y = 0.257 \text{ mol.m}^{-2}\text{s}^{-1}$.

IX.6 Calculate the mass transfer coefficient of benzene in gas phase for a plate stripping column. Benzene is stripped from coal tar at 130°C and 106 kPa . Ratio of free area of the plate to the cross-section of the column is 0.28 , hole diameters of the perforated plates is 10 mm . Vapour velocity given per column cross-section is 1.8 m.s^{-1} . Benzene concentration in gas is negligible. Assume, the physical properties of gaseous mixture are well approximated by pure steam properties.

For perforated plates, gas phase resistance was correlated using an equation:

$$\mathbf{Sh} = 0.94 \mathbf{Re}^{1.15} \mathbf{Sc}^{0.5} \mathbf{Ga}^{-0.114} \mathbf{S}^{0.74}$$

Where

Sherwood number	$\mathbf{Sh} = k_y d / c_G D_{AB}$
Reynolds number	$\mathbf{Re} = u_G d / \nu_G$
Schmidt number	$\mathbf{Sc} = \nu / D_{AB}$
Galileo number	$\mathbf{Ga} = g d^3 / \nu_G^2$

Characteristic dimension d is taken as the hole diameter, u_G is gas velocity in the holes.

Result: The mass transfer coefficient $k_y = 72.9 \text{ mol.m}^{-2}\text{s}^{-1}$.

IX.7 Calculate the mass transfer coefficient of benzene in liquid phase for a plate stripping column. Benzene is stripped from coal tar at 130°C and 106 kPa . Ratio of free area of the plate to the cross-section of the column is 0.28 , hole diameters of the perforated plates is 10 mm . Vapour velocity given per column cross-section is 1.8 m.s^{-1} . The tar density is 810 kg.m^{-3} , its viscosity is 0.75 mPa.s , surface tension 20.5 mN.m^{-1} , mean molar weight 253 g.mol^{-1} . Benzene diffusion coefficient in the coal tar at given conditions is $66.3 \times 10^{-7} \text{ m}^2\text{h}^{-1}$.

For perforated plates, liquid phase resistance was correlated using an equation:

$$\mathbf{Sh}_L = 5.42 \times 10^{-7} \mathbf{Re}_G^4 \mathbf{Sc}_L^{0.5} \mathbf{We}^{2.32} \mathbf{S}^{2.82}$$

Where

Sherwood number	$\mathbf{Sh}_L = k_x d / c_L D_{AC}$
Reynolds number	$\mathbf{Re}_G = u_G d / \nu_G$
Schmidt number	$\mathbf{Sc}_L = \nu_L / D_{AC}$
Weber number	$\mathbf{We} = \sigma / \rho_L g d^2$

Characteristic dimension d is taken as the hole diameter, u_G is gas velocity in the holes.

Result: The mass transfer coefficient $k_x = 2.41 \text{ mol.m}^{-2}\text{s}^{-1}$.

Tables

Many methods were suggested for calculation of diffusivity in gaseous mixtures at low pressure. A handbook² suggests following relation:

$$D_{AB} = \frac{0.01013 T^{1.75}}{P(v_A^{1/3} + v_B^{1/3})^2} \sqrt{\frac{1}{M_A} + \frac{1}{M_B}}$$

Where T is in K, P in Pa, M_A and M_B are in g/mol and v_A and v_B are gas parameters with dimension of molar volume. The diffusivity is then calculated in m².s⁻¹.

Values of the parameters v_i for some gases

gas	v	gas	v	gas	v
H ₂	7.07	CO	18.9	NH ₃	14.9
He	2.88	CO ₂	26.9	Air	20.1
N ₂	17.9	SO ₂	41.1	Cl ₂	37.7
O ₂	16.6	H ₂ O	12.7	Ar	16.1

If the values of the parameters v_i are unknown, they can be obtained using a contribution method.

Contributions of most common atoms and structure elements

Atom/structure	contribution	Atom/structure	contribution
C	16.5	Cl	19.5
H	1.98	S	17.0
O	5.48	Cycle	-20.2
N	5.69		

² Ried, Prausnitz and Poling : Properties of Gases and Liquids

X Packed absorption columns

X.1 A scrubber is used to process $1000 \text{ Nm}^3/\text{hr}$ of gas (measured at 0°C and pressure 101.3 kPa), which contains 100 g of CH_3OH per 1 Nm^3 . Pure water is used as solvent at the inlet. 98% of the CH_3OH entering the column gets absorbed into the water. At the outlet, water contains 67% of the maximum possible concentration.

The equilibrium relationship can be expressed by $Y = 1.15 X$.

The scrubber is filled with [ceramic Raschig rings](#) with diameter 25 mm . The wetting fraction of the packing is 60% . Mass transfer coefficient K_X is $0.5 \text{ kmol.m}^{-2}\text{h}^{-1}$.

The superficial velocity of the gas is 0.55 m.s^{-1} at 27°C and pressure 101.3 kPa .

Determine:

- Mass flow rate of water necessary for the process
- Diameter of the scrubber
- Height of the packing

Result: The mass flow rate of water is 1260 kg.h^{-1} , scruber diameter is 0.84 m , and packing height is 17.7 m .

X.2 Benzene is to be absorbed from gas into “washing oil” in a packed absorption column. The column operates at 30°C and 0.6 bar . 28.5 kmol.h^{-1} of gas containing $12.7 \text{ mol.}\%$ of benzene and 13 kmol.h^{-1} of the oil containing $1 \text{ mol.}\%$ of benzene is fed into the column. The gas at the outlet should contain at most $1.5 \text{ vol.}\%$. Assume ideal behaviour of the liquid phase and negligible saturated vapour pressure of the oil. Saturated vapour pressure of benzene at 30°C is 15.91 kPa . Calculate the number of transfer units (N_y).

Result: The number of transfer units is 3.74 .

X.3 SO_2 is to be removed from air by its absorption into water. The inlet gas contains $1.1 \text{ vol.}\%$ of SO_2 . Altogether, $35 \text{ Nm}^3.\text{hr}^{-1}$ of pure SO_2 is fed into the column, 98% of which is to be absorbed. The water at the outlet contains $0.31 \text{ wt.}\%$ of SO_2 . The superficial velocity of the gas is 0.45 m.s^{-1} . The column operates at 10°C and 100 kPa . The mass transfer coefficient $k_y a$ is $0.13 \text{ kmol.m}^{-3}\text{s}^{-1}$ and resistance to the mass transfer in liquid phase can be neglected. At these conditions, the equilibrium can be expressed by relation $p_{\text{SO}_2} = 3.3 \cdot 10^5 X_{\text{SO}_2}$, where p_{SO_2} is partial pressure of SO_2 in the gas and X_{SO_2} is relative mass fraction of SO_2 in the liquid.

Determine the height of packing.

Result: The packing height is 3.7 m (if mean logarithmic driving force is used).

X.4 Air containing 6 vol.% of NH₃ is to be purified from ammonia by its absorption into pure water at 20°C and 101,3 kPa. Superficial mass velocity of the gas at bottom of the column is 1.1 kg.m⁻²s⁻¹. The column packing is made of Berl saddles, 38mm diameter. The liquid flow density is 2.5 kg.m⁻²s⁻¹. Removal efficiency of ammonia should be 95%. Calculate the packing height assuming the liquid phase resistance is negligible and the equilibrium can be approximated by $Y = 0.153 X$, where Y and X are relative molar fractions of ammonia in gas and liquid, respectively. Assume the gas properties are unchanged by presence of ammonia.

For Rashig Rings, Pall rings, Berl saddles and spheres, the mass transfer coefficient in gas phase is given by equation:

$$\mathbf{Sh} = k \mathbf{Re}^{0.7} \mathbf{Sc}^{1/3} (a_t d_p)^{-2}$$

Where d_p is packing particle diameter, a_t is geometric area of the packing. For $d_p < 0.015$ m is $k = 2$; for $d_p > 0.015$ m the value of k is 5.23.

In this equation, the criteria are defined as follows

$$\mathbf{Sh} = k_y / c_L a_t D_{AC}$$

$$\mathbf{Re} = m_G / a_t S \eta_G$$

Result: The packing height is 0.53 m.

Numerical solution of absorption column for linear equilibrium

Packing height can be calculated using

$$h = \frac{n_{Li}}{K_x a_s S} \int_{x_{Ai}}^{x_{Ae}} \frac{dx}{x_A^* - x_A} = HTU_x NTU_x$$

$$h = \frac{n_{Gi}}{K_y a_s S} \int_{y_{Ae}}^{y_{Ai}} \frac{dx}{y_A - y_A^*} = HTU_y NTU_y$$

where HTU_x stands for height of transfer unit $HTU_x = \frac{n_{Li}}{K_x a_s S}$ and $HTU_y = \frac{n_{Gi}}{K_y a_s S}$

NTU_z is number of transfer units $NTU_x = \frac{x_{Ae} - x_{Ai}}{(x_A^* - x_A)_{LM}}$ and $NTU_y = \frac{y_{Ai} - y_{Ae}}{(y_A - y_A^*)_{LM}}$

Another way is to use absorption factor

$$\zeta_y = \frac{n_{Li}}{\psi_A n_{Gi}} \quad \text{or} \quad \zeta_x = \frac{1}{\zeta_y} = \frac{\psi_A n_{Gi}}{n_{Li}}$$

An absorption effect

$$\eta_x = \frac{x_{Ae} - x_{Ai}}{x_{Ae}^* - x_{Ai}} = \frac{\eta_y}{\zeta_y}$$

$$\eta_y = \frac{y_{Ai} - y_{Ae}}{y_{Ai} - y_{Ae}^*} = \frac{\eta_x}{\zeta_x}$$

Using these dimensionless variables, following relationship for number of transfer units (NTU) can be derived:

$$NTU_z = \frac{\zeta_z}{\zeta_z - 1} \ln \frac{1 - \frac{\eta_z}{\zeta_z}}{1 - \eta_z}$$

XI Membrane processes

Gas permeation

The basic flux across the membrane is exactly that for transport across a thin film. The flux j_A is proportional to the concentration difference

$$j_A = \frac{D}{l} (c_{A0} - c_{Al})$$

In which D is the diffusion coefficient, l is the membrane's thickness, c_{A0} is the concentration in the membrane on the feed side of the membrane, and c_{Al} is the concentration within the membrane on the permeate side of the membrane. The quantity D/l corresponds to a mass transfer coefficient. As the concentrations inside the membrane are not known, they must be replaced by known concentration in the feed and permeate using the partition coefficient (solubility)

$$j_A = \frac{DS}{l} (C_{A0} - C_{Al})$$

where S is the "solubility" between the membrane and the adjacent solution, defined by

$$c_A = S C_A$$

XI.1 Gas permeation is used to separate oxygen from nitrogen. Permeate should contain 95 mol.% of oxygen. Selectivity of the membrane is 5.4. Ratio of feed to permeate pressures $p_R/p_P = 4$ and cut is 0.6. Assuming ideal mixing on both sides of the membrane and ideal gas behaviour, determine the minimum oxygen content in the feed.

Result: The feed has to contain at least 89.9 mol.% of oxygen.

XI.2 Gas permeation is used to enrich oxygen content in air. Enriched air should contain 27 mol.% of oxygen. Selectivity of the membrane is 2.1. Reduced pressure $p_R = 2.86$. Assuming ideal mixing on both sides of the membrane. Determine the cut.

Result: The relative amount of permeate is 0.233.

Pervaporation

XI.3 By distillation at normal pressure, azeotropic mixture of ethanol and water is produced. This mixture is then separated by pervaporation using a PVA based membrane (polyvinylalcohol). The temperature in the PV module is 90°C and no additional heat is added. Retentate with water content 0.3wt.% is required. For water content in retentate below 10 wt.%, the flux of permeate is a linear function of the retentate water content and can be given as $J_P = 11.08 x_W$. Composition of permeate as a function of retentate composition was determined experimentally and is given in following table.

x_W	0	0.0025	0.005	0.0075	0.010	0.015	0.020
y_W	0	0.532	0.742	0.800	0.835	0.884	0.914
x_W	0.03	0.04	0.05	0.06	0.07	0.08	0.09
y_W	0.947	0.960	0.962	0.965	0.968	0.973	0.974

Determine: the cut
weight fraction of water in permeate
feed temperature
area of the membrane for processing $1300 \text{ kg}\cdot\text{h}^{-1}$ of feed.

Result: The cut is 0.073; water fraction in permeate is 0.565; feed temperature 125°C and membrane area 2860 m^2 .

Micro-filtration

XI.4 Aqueous suspension of china clay is to be concentrated using micro-filtration. Thickness of filtration membrane is $100 \mu\text{m}$ and it can be taken as ideal. The suspension feed contains 8 vol.% of solid particles. Concentrated suspension (retentate) should contain 15 vol.% of solid particles. The process is carried at 26°C and mean filtration pressure 0.7 MPa . By an experiment with pure water at 14°C and filtration pressure 0.36 MPa flux of filtrate was determined to be $416 \text{ L}\cdot\text{m}^{-2}\cdot\text{h}^{-1}$.

Determine the area of a filtration module to filtrate 125 m^3 of feed per day. What are going to be the flow rates of filtrate and concentrated suspension?

Determine the concentration of the solid particles in filtrate when a non-ideal membrane with identical permeability and rejection factor $R = 0.8$ is used. What are going to be the flow rates of filtrate and concentrated suspension and the filtration area?

Result: For ideal membrane with area 2.24 m^2 following amounts per day are produced: 58.3 m^3 of filtrate, 66.7 m^3 of concentrated suspension. For non-ideal membrane with area 2.80 m^2 following amounts per day are produced: 52.1 m^3 of concentrated suspension; 72.9 m^3 of filtrate containing 3 vol.% of solids.

Reverse osmosis

XI.5 Reverse osmosis unit is to be used to decrease NaCl content in sea water from $35 \text{ kg}\cdot\text{m}^{-3}$ down to $1.5 \text{ kg}\cdot\text{m}^{-3}$. Capacity of the unit should be 1000 m^3 of desalinated water per day. For the unit, membrane modules of area 3 m^2 , with a membrane which has a volumetric rejection factor 0.99 and permeate flux $54.2 \text{ L}\cdot\text{m}^{-2}\cdot\text{h}^{-1}$ at 16°C , 15bar pressure difference and NaCl concentration in retentate $1.5 \text{ kg}\cdot\text{m}^{-3}$. Desalination will be carried at pressure difference 200 bar. The osmotic pressure of NaCl is linear function of concentration being 0.372 MPa for solution containing $5 \text{ kg}\cdot\text{m}^{-3}$.

Determine: the number of membrane modules to be used.
the volumetric flow rate of the feed and relative fraction of permeate.
the concentration of NaCl in the retentate.

Result: 40 modules will have to be used. The flow rate of feed will be 1290 m^3 per day. Relative fraction of permeate 0.774 and NaCl concentration in retentate will be $150 \text{ kg}\cdot\text{m}^{-3}$.

XI.6 Reverse osmosis unit is to be used to clean waste solution of organic compounds. Volumetric flow rate of the waste solution is $26 \text{ m}^3\text{h}^{-1}$ and it contains 0.1 kmol.m^{-3} of organic compounds. Osmotic pressure of the solution is 0.45 MPa and a linear dependence of the osmotic pressure on the molar concentration can be assumed. The pressure difference in the membrane unit is 3 MPa . The cut fraction should be 0.4 . The properties of the membrane had been obtained experimentally: resistance of the membrane is $P_A/\delta_M = 0.5 \cdot 10^{-4} \text{ cm.MPa}^{-1}\text{s}^{-1}$ and the volumetric rejection factor is 0.9 . The module is ideally mixed on both sides of the membrane.

Determine: the volumetric flow rates of permeate and retentate,
concentration of the organic compounds in permeate and retentate,
area of the membrane module.

Result: The volumetric flow rate of the retentate is $15.6 \text{ m}^3\text{h}^{-1}$ and of permeate is $10.4 \text{ m}^3\text{h}^{-1}$.
The concentration of organics in retentate is $0.156 \text{ kmol.m}^{-3}$ and in permeate $0.0156 \text{ kmol.m}^{-3}$.
Area of the membrane necessary for the operation is 2450 m^2 .

XII Tables

Properties of liquid water

temperature [°C]	density [kg.m ⁻³]	dynamic viscosity [mPa.s]	heat capacity [kJ.kg ⁻¹ .K ⁻¹]	heat conductivity [W.m ⁻¹ .K ⁻¹]	Prandtl number [-]
0.01	999.78	1.792	4.229	0.5610	13.51
10	999.69	1.306	4.188	0.5800	9.435
20	998.19	1.002	4.183	0.5984	7.006
30	995.61	0.7977	4.183	0.6154	5.422
40	992.17	0.6532	4.182	0.6305	4.333
50	987.99	0.5470	4.182	0.6435	3.555
60	983.16	0.4665	4.183	0.6543	2.982
70	977.75	0.4040	4.187	0.6631	2.551
80	971.79	0.3544	4.194	0.6700	2.219
90	965.33	0.3145	4.204	0.6753	1.958
100	958.39	0.2818	4.217	0.6791	1.750
110	951.00		4.232	0.6817	
120	943.16		4.249	0.6832	
130	934.88		4.267	0.6837	
140	926.18		4.288	0.6833	
150	917.06		4.312	0.6821	

Properties of steam - water vapour

temperature [°C]	pressure [kPa]	heat of vaporization [kJ.kg ⁻¹]	heat capacity [kJ.kg ⁻¹ .K ⁻¹]	heat conductivity [W.m ⁻¹ .K ⁻¹]
0.01		2500	1.868	0.01707
20	2.31	2454	1.882	0.01823
40	7.36	2407	1.904	0.01960
60	19.9	2359	1.937	0.02118
80	47.4	2309	1.983	0.02301
100	101	2257	2.044	0.02509
120	199	2203	2.126	0.02746
140	361	2144	2.233	0.03014
160	619	2082	2.37	0.03312
180	1002	2014	2.56	0.03644
200	1554	1940	2.80	0.04010
220	2318	1859	3.11	0.04415

Equilibrium pressure, density and enthalpy of saturated water and water vapour as a function of temperature

t -temperature, p -pressure, ρ_l -liquid density, ρ_v -vapour density, h_l -liquid enthalpy, h_v -vapour enthalpy, Δh_{lv} -enthalpy of vapourization

t °C	p kPa	ρ_l kg m ⁻³	ρ_v kg m ⁻³	h_l kJ kg ⁻¹	h_v kJ kg ⁻¹	Δh_{lv} kJ kg ⁻¹
0.01	0.61173	999.78	0.004855	0.00	2500.5	2500.5
10	1.2281	999.69	0.009405	41.99	2518.9	2476.9
20	2.3388	998.19	0.017308	83.84	2537.2	2453.3
30	4.2455	995.61	0.030399	125.67	2555.3	2429.7
40	7.3814	992.17	0.05121	167.50	2573.4	2405.9
50	12.344	987.99	0.08308	209.33	2591.2	2381.9
60	19.932	983.16	0.13030	251.15	2260.8	2357.6
70	31.176	977.75	0.19823	293.01	2626.1	2333.1
80	47.373	971.79	0.29336	334.93	2643.1	2308.1
90	70.117	965.33	0.42343	376.93	2659.6	2282.7
100	101.32	958.39	0.5975	419.06	2675.7	2256.7
110	143.24	951.00	0.8260	461.34	2691.3	2229.9
120	198.48	943.16	1.1208	503.78	2706.2	2202.4
130	270.02	934.88	1.4954	546.41	2720.4	2174.0
140	361.19	926.18	1.9647	589.24	2733.8	2144.6
150	475.72	917.06	2.5454	632.32	2746.4	2114.1
160	617.66	907.50	3.2564	675.65	2758.0	2082.3
170	791.47	897.51	4.1181	719.28	2768.5	2049.2
180	1001.9	887.06	5.154	763.25	2777.8	2014.5
190	1254.2	876.15	6.390	807.60	2785.8	1978.2
200	1553.7	864.74	7.854	852.38	2792.5	1940.1
210	1906.2	852.82	9.581	897.66	2797.7	1900.0
220	2317.8	840.34	11.607	943.51	2801.3	1857.8
230	2795.1	827.25	13.976	990.00	2803.1	1813.1
240	3344.7	813.52	16.739	1037.24	2803.0	1765.7
250	3973.7	799.07	19.956	1085.32	2800.7	1715.4
260	4689.5	783.83	23.700	1134.38	2796.2	1661.9
270	5499.9	767.68	28.061	1184.57	2789.1	1604.6

Properties of dry air at normal pressure

Temperature [°C]	ρ [kg.m ⁻³]	c_p [kJ.kg ⁻¹ .K ⁻¹]	η [10 ⁻⁶ Pa.s]	λ [W.m ⁻¹ .K ⁻¹]
0	1.293	1.005	17.25	0.0241
20	1.205	1.005	18.18	0.0257
40	1.128	1.005	19.09	0.0273
60	1.06	1.005	19.98	0.0288
80	0.999	1.009	20.85	0.0303
100	0.946	1.009	21.71	0.0318
150	0.835	1.015	23.78	0.0354
200	0.746	1.026	25.75	0.0387
250	0.675	1.037	27.63	0.0420
300	0.616	1.047	29.42	0.0450
400	0.524	1.068	32.74	0.0508
500	0.456	1.093	35.75	0.0562

Properties of 30 wt.% solution of methanol

Temperature [°C]	10	20	30	40
density [kg.m ⁻³]	956	952	947	943
dynamic viscosity [mPa.s]	2.80	2.10	1.60	1.20

Properties of compounds in the state of ideal gas:

	Δh_f	a	b	c	d
NH3	-45720	27.31	0.02383	0.00001707	-1.185E-8
O2	0	28.11	-0.00368	0.00001746	-1.065E-8
NO	90430	29.35	-0.000938	0.000009747	-4.187E-9
N2	0	31.15	-0.01357	0.0000268	-1.168E-8
H2O	-242000	32.24	0.001924	0.00001055	-3.596E-9
CH3OH	-201300	21.15	0.07092	0.00002587	-2.852E-8
CH2O	-116000	23.48	0.03157	0.00002985	-2.300E-8
H2	0	27.14	0.009274	-0.00001381	7.645E-9
CO2	-393800	19.08	0.07344	-0.00005602	1.715E-8
S	0				
SO2	-297100	23.85	0.0699	-0.00004961	1.328E-8
SO3	-396000	19.21	0.1374	-0.00011760	3.70E-8

$$c_p = a + b \cdot T + c \cdot T^2 + d \cdot T^3 \quad T \text{ [K], } c_p \text{ [J / mol K]}$$

$$\Delta h_f \quad \text{heat of fusion in ideal gas state at 25°C, [J / mol]}$$

*Gas – liquid equilibrium data*CO₂ – aqueous diethanolamine solution (DEA) C_{DEA} – mol. Concentration of diethanolamine [$\text{mol}\cdot\text{dm}^{-3}$] X_{CO_2} – relative mol. Fraction of CO₂ to DEA p_{CO_2} – CO₂ partial pressure in gas

$t/^\circ\text{C}$	0		25		50	
$\frac{C_{DEA}}{10^3 \text{ mol m}^{-3}}$	$\frac{p_{CO_2}}{10^2 \text{ Pa}}$	X_{CO_2}	$\frac{p_{CO_2}}{10^2 \text{ Pa}}$	X_{CO_2}	$\frac{p_{CO_2}}{10^2 \text{ Pa}}$	X_{CO_2}
0,5	1000,8	1,119	976,3	0,987	888,5	0,883
	363,6	1,044	332,8	0,912	321,6	0,778
	362,1	1,035	130,5	0,797	94,4	0,588
	105,5	0,883	59,1	0,714	13,5	0,336
	14,7	0,741	14,7	0,551	-	-
2,0	1001,0	0,936	980,6	0,753	891,3	0,680
	362,8	0,837	971,9	0,813	323,0	0,562
	106,8	0,752	333,2	0,717	245,0	0,548
	15,3	0,604	132,4	0,633	94,7	0,489
	-	-	59,1	0,553	13,6	0,302
5,0	1006,7	0,762	988,0	0,661	909,8	0,562
	275,0	0,683	338,1	0,589	328,1	0,491
	105,9	0,638	59,7	0,506	95,7	0,414
	15,2	0,526	-	-	13,9	0,254
	-	-	-	-	-	-

Gas – liquid equilibrium for NH₃-H₂O X_{NH_3} – relative mass fraction of NH₃ in water, p_{NH_3} – partial pressure of NH₃ above saturated liquid

$t/^\circ\text{C}$	0	10	20	25	30	40	50	60
$X_{NH_3}/10^{-2}$	$p_{NH_3}/(10^2 \text{ Pa})$							
20	85,3	138,0	221,3	-	347	527	795	1112
15	56,9	93,5	152,0	-	239	364	540	777
10	33,5	55,7	92,8	-	147	223	329	481
7,5	23,6	39,9	66,7	-	106,3	160	239	348
5	14,8	25,5	42,3	-	68,0	102,0	153	220
4	-	21,5	33,2	-	53,5	81,1	121	172
3	-	15,1	24,3	31,3	39,5	60,0	89,5	126
2,5	-	-	20,0	25,9	32,5	-	-	103
2	-	-	16,0	20,4	25,7	-	-	81,3
1,6	-	-	-	16,0	20,4	-	-	64,9
1,2	-	-	-	12,1	15,3	-	-	48,4
1,0	-	-	-	9,87	-	-	-	40,3
0,5	-	-	-	4,53	-	-	-	-

Vapour-liquid equilibria

Legend:

 x, y mol fractions in liquid or vapour respectively $t(x)$ boiling temperature for mol based fraction in liquid X, Y mass fractions in liquid or vapour respectively $t(X)$ boiling temperature for mass based fraction in liquid

ethanol - n-butanol				
$x.X$	y	$t(x)$	Y	$t(X)$
0.00	0.0000	117.45	0.0000	117.45
0.01	0.0420	116.55	0.0421	116.02
0.03	0.1190	114.82	0.1196	113.39
0.05	0.1879	113.18	0.1891	111.03
0.10	0.3311	109.46	0.3337	106.08
0.15	0.4429	106.18	0.4459	102.12
0.20	0.5319	103.26	0.5348	98.87
0.25	0.6042	100.64	0.6065	96.16
0.30	0.6637	98.27	0.6654	93.84
0.35	0.7135	96.12	0.7145	91.84
0.40	0.7556	94.15	0.7561	90.09
0.45	0.7917	92.33	0.7918	88.53
0.50	0.8230	90.65	0.8227	87.14
0.55	0.8503	89.08	0.8499	85.89
0.60	0.8745	87.61	0.8739	84.76
0.65	0.8959	86.24	0.8954	83.72
0.70	0.9152	84.93	0.9147	82.77
0.75	0.9325	83.70	0.9321	81.89
0.80	0.9484	82.53	0.9481	81.07
0.85	0.9628	81.41	0.9626	80.31
0.90	0.9762	80.33	0.9761	79.60
0.95	0.9885	79.30	0.9885	78.93
0.97	0.9932	78.89	0.9932	78.67
0.99	0.9978	78.50	0.9978	78.42
1.00	1.0000	78.30	1.0000	78.30

heptane - octane				
$x.X$	y	$t(x)$	Y	$t(X)$
0.00	0.0000	125.60	0.0000	125.60
0.01	0.0232	125.12	0.0232	125.05
0.03	0.0678	124.17	0.0678	123.99
0.05	0.1101	123.26	0.1101	122.96
0.10	0.2068	121.09	0.2068	120.57
0.15	0.2926	119.06	0.2925	118.38
0.20	0.3693	117.16	0.3692	116.36
0.25	0.4382	115.37	0.4382	114.49
0.30	0.5006	113.67	0.5006	112.75
0.35	0.5574	112.05	0.5575	111.11
0.40	0.6094	110.52	0.6095	109.57
0.45	0.6571	109.05	0.6572	108.12
0.50	0.7010	107.63	0.7012	106.74
0.55	0.7416	106.28	0.7418	105.43
0.60	0.7792	104.97	0.7794	104.18
0.65	0.8141	103.71	0.8143	102.99
0.70	0.8466	102.49	0.8468	101.85
0.75	0.8768	101.31	0.8770	100.76
0.80	0.9050	100.17	0.9051	99.72
0.85	0.9313	99.06	0.9314	98.71
0.90	0.9558	97.99	0.9558	97.75
0.95	0.9787	96.94	0.9787	96.82
0.97	0.9874	96.54	0.9874	96.46
0.99	0.9959	96.13	0.9959	96.11
1.00	1.0000	95.93	1.0000	95.93

Methanol - Water				
$x.X$	y	$t(x)$	Y	$t(X)$
0.00	0.0000	100.00	0.0000	100.00
0.01	0.0763	98.07	0.0770	98.88
0.03	0.1948	94.84	0.1994	96.84
0.05	0.2822	92.25	0.2922	95.03
0.10	0.4257	87.52	0.4486	91.26
0.15	0.5144	84.28	0.5461	88.30
0.20	0.5767	81.84	0.6134	85.88
0.25	0.6247	79.90	0.6633	83.85
0.30	0.6640	78.26	0.7025	82.10
0.35	0.6979	76.83	0.7348	80.54
0.40	0.7282	75.54	0.7624	79.12
0.45	0.7558	74.36	0.7868	77.81
0.50	0.7817	73.26	0.8089	76.56
0.55	0.8062	72.23	0.8295	75.36
0.60	0.8296	71.25	0.8491	74.19
0.65	0.8523	70.31	0.8680	73.04
0.70	0.8744	69.40	0.8864	71.88
0.75	0.8960	68.53	0.9048	70.72

0.80	0.9173	67.68	0.9231	69.54
0.85	0.9382	66.85	0.9417	68.34
0.90	0.9590	66.05	0.9606	67.10
0.95	0.9796	65.27	0.9800	65.82
0.97	0.9878	64.96	0.9879	65.30
0.99	0.9959	64.65	0.9959	64.77
1.00	1.0000	64.50	1.0000	64.50

*Liquid-liquid equilibria***Equilibrium data for immiscible solvent systems, expressed in relative weight fractions of acetone:**

Water – Acetone – Toluene	
aqueous phase	organic phase
0.0059	0.0036
0.0175	0.0112
0.0277	0.0188
0.0354	0.0251
0.0438	0.0325
0.0553	0.0426
0.0652	0.0530
0.0782	0.0636
0.0883	0.0727
0.1038	0.0883

Water – Acetone - o-Xylene	
aqueous phase	organic phase
0.0092	0.0060
0.0183	0.0120
0.0370	0.0236
0.0547	0.0382
0.0865	0.0660
0.0989	0.0735
0.1891	0.1583
0.2821	0.2508

water – acetic acid – diethyl ether at 25°C

x – weight fraction

Binodal curve					
$\frac{x}{10^{-2}}$					
water	diethyl ether	acetic acid	water	diethyl ether	acetic acid
2,3	97,7	0	22,3	49,3	28,4
5,5	85,2	9,3	33,2	36,5	30,3
9,4	73,7	16,9	42,9	26,6	30,5
11,9	67,8	20,3	54,9	16,8	28,3
17,5	56,7	25,8	66,0	11,3	22,7
78,7	7,6	13,7	93,3	6,7	0
86,6	7,1	6,3			

Equilibrium phase composition					
$\frac{x}{10^{-2}}$					
Diethyl ether phase			aqueous phase		
water	diethyl ether	acetic acid	water	diethyl ether	acetic acid
23,6	47,7	28,7	55,7	16,4	27,9
15,1	61,3	23,6	65,0	11,9	23,1
10,4	71,5	18,1	72,1	9,5	18,4
7,2	80,3	12,5	78,2	8,0	13,8
5,0	87,7	7,3	84,0	7,2	8,8
3,6	92,6	3,8	88,0	6,9	5,1

aniline - cyclohexane - heptane at 25°C

x - weight fraction

Binodal curve					
$\frac{x}{10^{-2}}$					
aniline	heptane	cyclohexane	aniline	heptane	cyclohexane
6,60	93,20	0	93,50	6,50	0
8,20	71,90	19,90	86,00	4,00	10,00
8,14	69,96	21,90	79,80	2,40	17,80
10,30	46,50	43,20	73,80	0	26,20
12,55	27,94	59,51			
15,22	14,31	70,47			
16,50	0	83,50			
Equilibrium phase composition					
$\frac{x}{10^{-2}}$					
Cyclohexane phase			Aniline phase		
aniline	heptane	cyclohexane	aniline	heptane	cyclohexane
13,0	25,6	61,4	83,9	3,4	12,7
9,3	59,3	31,4	90,5	5,4	4,1
10,4	47,8	41,8	88,9	4,9	6,2
8,1	73,5	18,4	92,5	6,0	1,5

water - acetone - trichloroethane at 25°C

x - weight fraction

Binodal curve					
$\frac{x}{10^{-2}}$					
water	C ₂ H ₃ Cl ₃	acetone	water	C ₂ H ₃ Cl ₃	acetone
0,11	99,89	0,00	16,63	23,20	60,17
0,26	94,73	5,01	19,31	20,71	59,98
0,36	90,11	9,53	26,28	15,35	58,33
0,59	84,65	14,76	34,96	10,00	55,04
0,76	79,58	19,66	35,38	9,63	54,99
0,99	75,72	23,29	41,35	6,77	51,88
1,43	70,36	28,21	48,47	4,35	47,18
1,44	67,52	31,04	55,97	2,18	41,85
1,87	64,17	33,96	61,11	1,72	37,17
2,11	60,06	37,83	66,58	1,17	32,25
2,98	54,88	42,14	71,80	1,02	27,18
4,01	48,78	47,21	74,54	0,92	24,54
5,0	43,88	51,12	80,40	0,78	18,82
6,84	38,31	54,85	84,94	0,70	14,36
9,78	31,67	58,55	87,63	0,65	11,72
13,35	26,39	60,26	94,66	0,52	4,82
15,37	24,04	60,59	99,56	0,44	0,00

Equilibrium phase composition					
$\frac{x}{10^{-2}}$					
Aqueous phase			Organic phase		
water	C ₂ H ₃ Cl ₃	acetone	water	C ₂ H ₃ Cl ₃	Acetone
93,52	0,52	5,95	0,32	90,93	8,75
92,95	0,54	6,51	0,40	89,32	10,28
85,35	0,68	13,97	0,90	78,32	20,78
82,23	0,73	17,04	1,10	73,76	25,14
80,16	0,79	19,05	1,33	71,01	27,66
73,00	1,00	26,00	2,09	60,85	37,06
72,06	1,02	26,92	2,27	59,21	38,52
71,33	1,04	27,63	2,40	58,21	39,39
69,35	1,11	29,54	2,85	55,48	41,67
67,95	1,17	30,88	3,11	53,92	42,97
62,67	1,60	35,73	4,26	47,53	48,21
57,00	2,10	40,90	6,05	40,00	53,95
50,20	3,75	46,05	8,90	33,70	57,40
41,70	6,52	51,78	13,40	26,26	60,34
	Critical point		27,40	14,60	58,09

water - methanol - trichloroethylene at 20°C*x* - mol fraction

Binodal curve					
$\frac{x}{10^{-2}}$					
water	trichloroethylene	methanol	water	trichloroethylene	methanol
94,94	0,027	5,02	22,71	17,64	59,67
89,9	0,041	10,05	19,98	21,6	58,45
84,73	0,067	15,22	18,63	24,52	56,80
79,92	0,106	19,95	17,11	27,96	54,91
74,86	0,154	24,96	15,11	32,85	52,05
69,83	0,251	29,81	13,34	36,98	49,68
70,75	0,220	29,10	11,72	41,70	46,6
58,6	0,88	40,6	9,76	47,90	42,35
50,2	1,81	48,0	7,60	55,00	37,24
46,25	2,62	51,14	5,10	65,04	29,85
40,13	4,48	55,40	3,575	71,90	24,62
33,01	7,87	59,13	2,03	81,68	16,30
29,12	10,37	60,50			

Equilibrium phase composition					
$\frac{x}{10^{-2}}$					
Aqueous phase			Organic phase		
water	trichloroethylene	methanol	water	trichloroethylene	methanol
82,3	0,15	17,55	0,1	98,6	1,3
69,8	0,3	29,9	0,2	97,25	2,55
60,2	0,7	39,1	0,3	95,6	4,1
50,0	1,7	48,3	0,6	92,6	6,8
41,9	3,7	54,4	1,15	88,75	10,1
35,1	6,6	58,3	1,7	83,3	15,0
29,8	9,9	60,3	2,6	77,6	19,8
26,2	13,1	60,7	3,7	71,3	25,0
21,4	19,35	59,25	5,7	62,25	32,05
17,9	26,35	55,75	8,55	51,55	39,9
15,7	30,95	53,35	10,25	46,15	43,6
15,05	32,85	52,1	11,3	43,15	45,65

Integral enthalpy of dissolution at 25°C, in kJ.kg^{-1}

	w												
	0	0.01	0.02	0.05	0.1	0.15	0.2	0.25	0.3	0.35	0.4	0.45	0.5
NaOH	-1072	-1055	-1055	-1065	-1071	-1070	-1060	-999	-954	-871	-762	-670	-608
KOH	-985	-974	-969	-968	-964	-958	-947	-932	-912	-882	-844	-806	-754
KCl	231												
NaCl	66.6	73	72	65.5	55	45.3	37	32.5					
NH ₄ Cl	275	285	286	286	286	284	284	282	281				
CaCl ₂	-748	-729	-727	-719	-711	-704							
KNO ₃	121	119											
NH ₄ NO ₃	322												

Increase of boiling point in solutions of electrolytes