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FOURTH EDITION

COULSON & RICHARDSON'S CHEMICAL ENGINEERING SERIES

Chemical Engineering Design

VOLUME 6

Solution Manual

Problem 1.1

(i) 1 ft = 0.305 m

(ii) 1 lb_m = 0.454 kg

(iii) 1 lb_f = 4.45 N

(iv) 1 HP = 746 W

(v) 1 psi = 6.9 kN m^{-2}

(vi) 1 lb ft s^{-1} = 1.49 N s m^{-2}

(vii) 1 poise = 0.1 N s m^{-2}

(viii) 1 Btu = 1.056 kJ

(ix) 1 CHU = 2.79 kJ

(x) 1 Btu ft^{-2} h^{-1} $^oF^{-1}$ = 5.678 W m^{-2} K^{-1}

Examples:

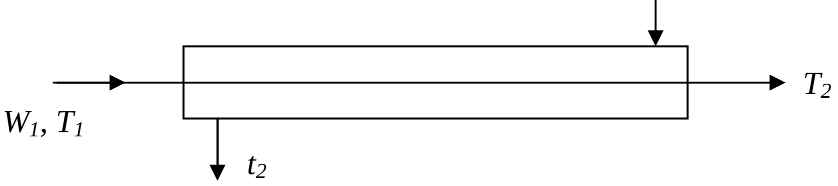
(viii) 1 Btu = 1 lb_m of water through 1 ^oF
= 453.6 g through 0.556 ^oC
= 252.2 cal
= (252.2)(4.1868)
= 1055.918 J = 1.056 kJ

$$\text{(x) } 1\ \text{Btu ft}^{-2}\ \text{h}^{-1}\ {}^o\text{F}^{-1} = \left\{1\,\text{Btu x}\,1.056\ \text{x}\,10^3\left(\frac{\text{J}}{\text{Btu}}\right)\right\}\text{x}\left\{1\ \text{ft x 12 x 25.4 x}\ 10^{-3}\left(\frac{\text{m}}{\text{ft}}\right)\right\}^{-2}$$

$$\text{x}\left\{1\,\text{h x 3600}\left(\frac{\text{s}}{\text{h}}\right)\right\}^{-1}\text{x}\left\{1\ {}^o\text{F x 0.556}\left(\frac{{}^o\text{C}}{{}^o\text{F}}\right)\right\}^{-1}$$

= 5.678 W m^{-2} $^oC^{-1}$
= 5.678 W m^{-2} K^{-1}

Problem 1.2



Variables, M:

1. Duty, heat transferred, Q
2. Exchanger area, A
3. Overall coefficient, U
4. Hot-side flow-rate, W_1
5. Cold-side flow-rate, W_2
6. Hot-side inlet temperature, T_1
7. Hot-side outlet temperature, T_2
8. Cold-side inlet temperature, t_1
9. Cold-side outlet temperature, t_2

Total variables = 9

Design relationships, N:

1. General equation for heat transfer across a surface

   $Q = UA\Delta T_m$ (Equation 12.1)

   Where ΔT_m is the LMTD given by equation (12.4)
2. Hot stream heat capacity $Q = W_1 C_p (T_1 - T_2)$
3. Cold stream heat capacity $Q = W_2 C_p (t_2 - t_1)$
4. U is a function of the stream flow-rates and temperatures (see Chapter 12)

Total design relationships = 4

So, degrees of freedom = $M - N = 9 - 4 = 5$

Problem 1.3

Number of components, $C = 3$

Degrees of freedom for a process stream = $C + 2$ (see Page 17)

Variables:

Streams	$4(C + 2)$
Separator pressure	1
Separator temperature	1
Total	$4C + 10$

Relationships:

Material balances	C
v-l-e relationships	C
l-l-e relationships	C
Equilibrium relationships	6
Total	$3C + 6$

Degrees of freedom = $(4C + 10) - (3C + 6) = C + 4$

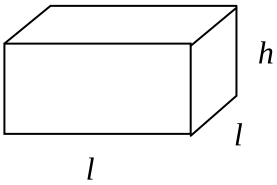
For $C = 3$, degrees of freedom = 7

The feed stream conditions are fixed which fixes $C + 2$ variables and so the design variables to be decided = 7 – 5 = 2.

Choose temperature and pressure.

Note: temperature and pressure taken as the same for all streams.

Problem 1.4



Volume = $l^2 \times h = 8 \text{ m}^3$

(i) *Open Top*

$$\text{Area of plate} = l^2 + 4lh$$

$$= l^2 + 4l \times 8l^{-2}$$

$$\text{Objective function} = l^2 + 32l^{-1}$$

Differentiate and equate to zero:

$$0 = 2l - 3l^{-2}$$

$$l = \sqrt[3]{16} = 2.52 \text{ m} \qquad \text{i.e. } h = \frac{l}{2}$$

(ii) *Closed Top*

The minimum area will obviusly be given by a cube, $l = h$

Proof:

Area of plate = $2l^2 + 4lh$

Objective function = $2l^2 + 32l^{-1}$

Differentiate and equate to zero:

$$0 = 4l - 3l^{-2}$$

$$l = \sqrt[3]{8} = 2 \text{ m}$$

$$h = \frac{8}{2^2} = 2 \text{ m}$$

Problems 1.5 and 1.6

Insulation problem, spread-sheet solution

All calculations are peformed per m^2 area

Heat loss = (U)(temp. diff.)(sec. in a year)

Savings = (heat saved)(cost of fuel)

Insulation Costs = (thickness)(cost per cu. m)(capital charge)

Thickness (mm)	U ($Wm^{-2}C^{-1}$)	Heat Loss (MJ)	Increment Savings (£)	Extra Cost Insulation (£)	
0	2.00	345.60	20.74		
25	0.90	155.52	11.40	0.26	
50	0.70	120.96	2.07	0.26	
100	0.30	51.84	4.15	0.53	(Optimum)
150	0.25	43.20	0.52	0.53	
200	0.20	34.56	0.52	0.53	
250	0.15	25.92	0.52	0.53	

Data: cost of fuel 0.6p/MJ

av. temp. diff. 10^{o}C

200 heating days per year

cost of insulation £70/m^3

capital charges 15% per year

American version:

Thickness (mm)	U ($Wm^{-2}C^{-1}$)	Heat Loss (MJ/yr)	Increment Savings ($/$m^2$)	Extra Cost Insulation ($/$m^2$)	
0	2.00	518.40	45.66		
25	0.90	233.28	25.66	0.6	
50	0.70	181.44	4.66	0.6	
100	0.30	77.76	9.33	1.2	(Optimum)
150	0.25	64.80	1.17	1.2	
200	0.20	51.84	1.17	1.2	
250	0.15	38.88	1.17	1.2	

Data: cost of fuel 0.6 cents/MJ
av. temp. diff. 12°C
250 heating days per year
cost of insulation $120/$m^3$
capital charges 20% per year

Problem 1.7

The optimum shape will be that having the lowest surface to volume ratio.
A sphere would be impractical to live in an so a hemisphere would be used.
The Inuit build their snow igloos in a roughly hemispherical shape.
Another factor that determines the shape of an igloo is the method of construction. Any cross-section is in the shape of an arch; the optimum shape to use for a material that is weak in tension but strong in compression.

Problem 1.8

1. THE NEED
   Define the objective:
   a) purging with inert gas, as requested by the Chief Engineer
   b) safety on shut down

2. DATA

Look at the process, operation, units, flammability of materials, flash points and explosive limits.

Read the report of the incident at the similar plat, if available. Search literature for other similar incidents.

Visit sites and discuss the problem and solutions.

Determine volume and rate of purging needed.

Collect data on possible purging systems. Discuss with vendors of such systems.

3. GENERATION OF POSSIBLE DESIGNS

Types of purge gase used: Argon, helium, combustion gases (CO_2 + H_2O), nitrogen and steam.

Need to consider: cost, availability, reliability, effectiveness.

Helium and argon are rejected on grounds of costs and need not be considered.

a) Combustion gases: widely used for purging, use oil or natural gas, equipment readily available: consider.
b) Nitrogen: used in process industry, available as liquid in tankers or generated on site: consider.
c) Steam: used for small vessels but unlikely to be suitable for a plant of this size: reject.

4. EVALUATION:

Compare combustion gases versus nitrogen.

- *Cost*

Cost of nitrogen (Table 6.5) 6p/m^3

Cost of combustion gases will depend on the fuel used. Calculations are based on natural gas (methane).

$$2CH_4 + 3O_2 + (3x4)N_2 \rightarrow 2CO_2 + 4H_2O + 12N_2$$

So, 1 m^3 of methane produces 7 m^3 of inert combustion gases (water will be condensed).

Cost of natural gas (Table 6.5) 0.4p/MJ. Typical calorific value is 40 MJ/m^3.

Therefore, cost per m^3 = 0.4 x 40 = 16p.

Cost per m^3 of inert gases = 16/7 = 2.3p.

So, the use of natural gas to generate inert gas for purging could be significantly cheaper than purchasing nitrogen. The cost of the generation equipment is not likely to be high.

- *Availability*

  Natural gas and nitrogen should be readily available, unless the site is remote.

- *Reliability*

  Nitrogen, from storage, is likely to be more reliable than the generation of the purge gas by combustion. The excess air in combustion needs to be strictly controlled.

- *Effectiveness*

  Nitrogen will be more effective than combustion gases. Combustion gases will always contain a small amount of oxygen. In addition, the combustion gases will need to be dried thoroughly and compressed.

5. FINAL DESIGN RECOMMENDATION

Use nitrogen for the large scale purging of hazardous process plant.

Compare the economics of generation on site with the purchase of liquid nitrogen. Generation on site would use gaseous storage, under pressure. Purchase would use liquid storage and vapourisation.

Solution 2.1

Basis for calculation: 100 kmol dry gas

Reactions:

$$CO + 0.5O_2 \rightarrow CO_2$$

$$H_2 + 0.5O_2 \rightarrow H_2O$$

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

$$C_2H_6 + 3.5O_2 \rightarrow 2CO_2 + 3H_2O$$

$$C_6H_6 + 7.5O_2 \rightarrow 6CO_2 + 3H_2O$$

	REACTANTS		PRODUCTS		
	Nat. Gas	O_2	CO_2	H_2O	N_2
CO_2	4		4		
CO	16	8	16		
H_2	50	25		50	
CH_4	15	30	15	30	
C_2H_6	3	10.5	6	9	
C_6H_6	2	15	12	6	
N_2	10				10
Totals	100	88.5	53	95	10

If Air is N_2:O_2 = 79:21

N_2 with combustion air = 88.5 x 79/21 = 332.9 kmol

Excess O_2 = 88.5 x 0.2 = 17.7 kmol

Excess N_2 =17.7 x 79/21 = 66.6 kmol

Total = 417.2 kmol

(i) Air for combustion = 417.2 + 88.5 = 505.7 kmol

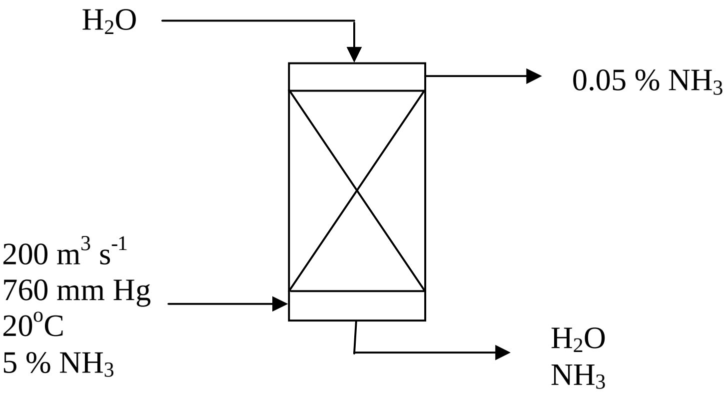
(ii) Flue Gas produced = 53 + 95 + 10 + 417.2 = 575.2 kmol

(iii) Flue Gas analysis (dry basis):

N_2	409.5 kmol	85.3 mol %
CO_2	53.0 kmol	11.0 mol %
O_2	17.7 kmol	3.7 mol %
	480.2 kmol	100.0 mol %

Solution 2.2

Use air as the tie substance – not absorbed.



Partial volume of air = 200(1 - 0.05) = 190 $m^3\ s^{-1}$

Let the volume of NH_3 leaving the column be x, then:

$$\frac{0.05}{100} = \frac{x}{190 + x}$$

$0.05(190 + x) = 100x$

$$x = \frac{9.5}{(100 - 0.05)} = 0.0950\ \text{m}^3\ \text{s}^{-1}$$

(a) The volume of NH_3 adsorbed = (200)(0.05) – 0.0950

= 9.905 $m^3\ s^{-1}$

If 1 kmol of gas occupies 22.4 m^3 at 760 mm Hg and 0°C,

$$\text{Molar Flow} = \left(\frac{9.905}{22.4}\right)\frac{273}{(273 + 20)} = 0.412\ \text{kmol s}^{-1}$$

Mass Flow = (0.412)(17) = 7.00 kg s^{-1}

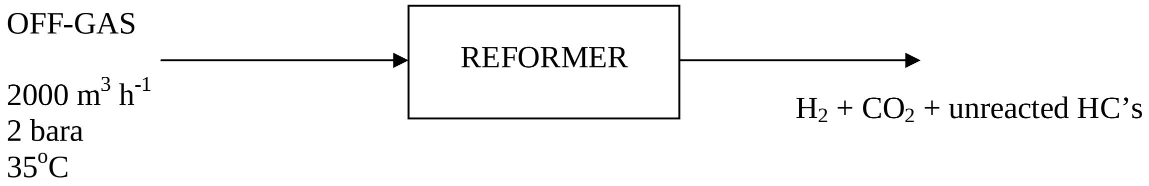
(b) Flow rate of gas leaving column = 190 + 0.0950 = 190.1 $m^3\ s^{-1}$

(c) Let the water flow rate be W, then:

$$\frac{1}{100} = \frac{7.00}{W + 7.00}$$

W = 700 – 7 = 693 kg s^{-1}

Solution 2.3



At low pressures vol% = mol%

(a) Basis: 1 kmol of off-gas

Component	mol%	M. M.		mass (kg)
CH_4	77.5	16		12.40
C_2H_6	9.5	30		2.85
C_3H_8	8.5	44		3.74
C_4H_{10}	4.5	58		2.61
			Σ	21.60

So the average molecular mass = 21.6 kg kmol^{-1}

(b) At STP, 1 kmol occupies 22.4 m^3

$$\text{Flow rate of gas feed} = \left(\frac{2000}{22.4}\right)\left(\frac{2\text{ x }10^5}{1.013\text{ x }10^5}\right)\frac{273}{(273+35)} = 156.248\text{ kmol h}^{-1}$$

Mass flow rate = (156.248)(21.60) = 3375 kg h^{-1}

(c) Basis: 100 kmol of feed

Reaction (1): $C_nH_{2n+2} + n(H_2O) \rightarrow n(CO) + (2n + 1)H_2$

Component	*n*	Amount		CO	H_2
CH_4	1	77.5		77.5	232.5
C_2H_6	2	9.5		19.0	47.5
C_3H_8	3	8.5		25.5	59.5
C_4H_{10}	4	4.5		18.0	40.5
			Σ	140.0	380.0

If the conversion is 96%, then:

H_2 produced = (380.0)(0.96) = 364.8 kmol

CO produced = (140.0)(0.96) = 134.4 kmol

Reaction (2): $CO + H_2O \rightarrow CO_2 + H_2$

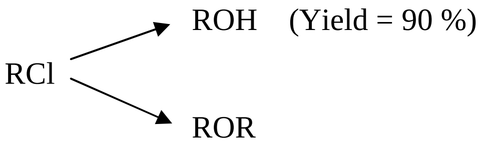
If the conversion is 92%, then: H_2 from CO = (134.4)(0.92) = 123.65 kmol

Total H_2 produced = 364.8 + 123.65 = 488.45 kmol/100 kmol feed

If the gas feed flow rate = 156.25 kmol h^{-1}, then

$$H_2 \text{ produced} = 156.25\left(\frac{488.45}{100}\right) = 763.20 \text{ kmol h}^{-1} \equiv (763.2)(2) = 1526 \text{ kg h}^{-1}$$

Solution 2.4



(Conversion = 97 %)

Basis: 1000 kg RCl feed

Relative molecular masses:

$CH_2{=}CH{-}CH_2Cl$	76.5
$CH_2{=}CH{-}CH_2OH$	58.0
$(CH_2{=}CH{-}CH_2)_2O$	98.0

RCl feed $= \dfrac{1000}{76.5}$ = 13.072 kmol

RCl converted = (13.072)(0.97) = 12.68 kmol

ROH produced = (12.68)(0.9) = 11.41 kmol

ROR produced = 12.68 – 11.41 = 1.27 kmol

Mass of allyl-alcohol produced = (11.41)(58.0) = 661.8 kg

Mass of di-ally ether produced = (1.27)(98.0) = 124.5 kg

Solution 2.5

Basis: 100 kmol nitrobenzene feed.

The conversion of nitrobenzene is 96% and so 100(1 - 0.96) = 4 kmol are unreacted.

The yield to aniline is 95% and so aniline produced = (100)(0.95) = 95 kmol

Therefore, the balance is to cyclo-hexalymine = 96 – 95 = 1 kmol

From the reaction equations:

$$C_6H_5NO_2 \ + \ 3H_2 \ \rightarrow \ C_6H_5NH_2 \ + \ 2H_2O$$

1 mol of aniline requires 3 mol of H_2

$$C_6H_5NO_2 \ + \ 6H_2 \ \rightarrow \ C_6H_{11}NH_2 \ + \ 2H_2O$$

1 mol of cyclo-hexalymine requires 6 mol of H_2

Therefore, H_2 required for the reactions = (95)(3) + (1)(6) = 291 kmol

A purge must be taken from the recycle stream to maintain the inerts below 5%. At steady-state conditions:

Flow of inerts in fresh H_2 feed = Loss of inerts from purge stream

Let the purge flow be x kmol and the purge composition be 5% inerts.

Fresh H_2 feed = H_2 reacted + H_2 lost in purge

= 291 + (1 – 0.05)x

Inerts in the feed at 0.005 mol fraction (0.5%) $= (291 + 0.95x)\dfrac{0.005}{1 - 0.005}$

$= 1.462 + 4.774 \times 10^{-3}x$

Inerts lost in purge = $0.05x$

So, equating these quantities: $0.05x = 1.462 + 4.774 \times 10^{-3}x$

Therefore: $x = 32.33$ kmol

The purge rate is 32.33 kmol per 100 kmol nitrobenzene feed.

H_2 lost in the purge = 32.33(1 – 0.05) = 30.71 kmol

Total H_2 feed = 291 + 30.71 = 321.71 kmol

Therefore: Total feed including inerts $= \dfrac{321.71}{1 - 0.005}$ = 323.33 kmol

(c) *Composition at the reactor outlet*:

Stoichiometric H_2 for aniline = 285 kmol

H_2 feed to the reactor = (285)(3) = 855 kmol

Fresh feed H_2 = 323.33 and so Recycle H_2 = 855 – 323.33 = 531.67 kmol

Inerts in Fresh Feed = (323.33)(0.005) = 1.617 kmol

Inerts in Recycle (at 5%) $= 536.08\left(\dfrac{0.05}{1 - 0.05}\right)$ = 27.983 kmol

Therefore, total inerts = 1.617 + 27.983 = 29.600 kmol

Aniline produced = 95 kmol

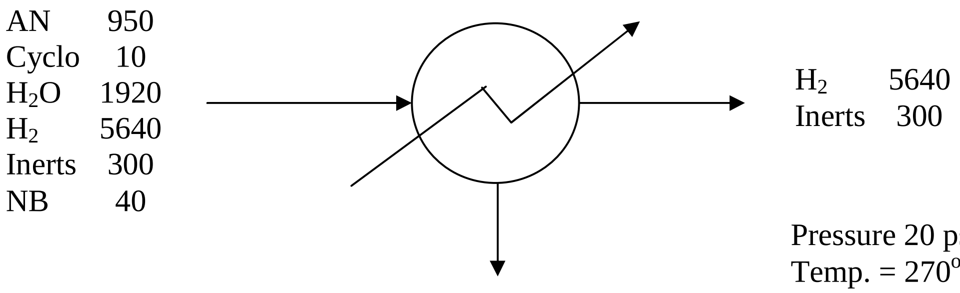
Cyclo-hexalymine produced = 1 kmol

If 291 kmol of H_2 are reacted, then H_2 leaving the reactor = 855 – 291 = 564 kmol

H_2O produced = (95)(2) + (1)(2) = 192 kmol

Composition:	kmol	mol %
Aniline	95	10.73
Cyclo-hexalymine	1	0.11
H_2O	192	21.68
H_2	564	63.69
Inerts	29.60	3.34
Nitrobenzene	4	0.45
	885.6	100.00

Solution 2.6



Pressure 20 psig = 1.38 barg
Temp. = 270°C

Assumptions: H_2 and inerts are not condensed within the condenser.

Temp. of the gas at the condenser outlet = 50°C and return the cooling water at 30°C (20°C temp. difference).

Antoine coefficients:	Aniline	16.6748, 3857.52, -73.15
	Nitrobenzene	16.1484, 4032.66, -71.81
	H_2O	18.3036, 3816.44, -46.13

Vapour pressures at 50^oC:

H_2O: $$\ln(P^o) = 18.3036 - \frac{3816.44}{323 - 46.13}$$

P^o = 91.78 mm Hg = 0.122 bar   (From Steam Tables = 0.123 bar)

Aniline: $$\ln(P^o) = 16.6748 - \frac{3857.52}{323 - 73.15}$$

P^o = 3.44 mm Hg = 0.00459 bar

Nitrobenzene: $$\ln(P^o) = 16.1484 - \frac{4032.66}{323 - 71.81}$$

P^o = 1.10 mm Hg = 0.00147 bar

NB. The cyclo-hexalymine is ignored because it is present in such a small quantity.

$$\text{Mol fraction} = \frac{\text{partial pressure}}{\text{total pressure}}$$

If the total pressure is 2.38 bara

$$H_2O = \frac{0.122}{2.38} = 0.0513 = 5.13\ \%$$

$$AN = \frac{0.00459}{2.38} = 0.0019 = 0.19\ \%$$

$$NB = \frac{0.00147}{2.38} = 0.00062 = 0.06\ \%$$

Total   5.38 %

Take H_2 and the inerts as tie materials.

Flow (H_2 and inerts) = 5640 + 300 = 5940 kmol

Mol fraction (H_2 and inerts) = 100 – 5.38 = 94.62 %

$$\text{Flow of other components} = \left(\frac{\text{mol fraction other}}{\text{mol fraction (H}_2 + \text{inerts)}}\right)\text{flow (H}_2 + \text{inerts)}$$

$$H_2O = \frac{5.13}{94.53} \text{ x } 5940 = 322.0 \text{ kmol}$$

$$AN = \frac{0.19}{94.53} \text{ x } 5940 = 11.9 \text{ kmol}$$

$$NB = \frac{0.06}{94.53} \text{ x } 5940 = 3.8 \text{ kmol}$$

Composition of the gas stream (recycle):

	kmol	vol %
H_2	5640	89.84
Inerts	300	4.78
H_2O	322.0	5.13
AN	11.9	0.19
NB	3.8	0.06
Cycl.	Trace	--
Total	6277.7	100.00

Composition of the liquid phase:

Liquid Flow = Flow In – Flow in Gas Phase

		kmol	kg	vol %	w/v %
H_2		0	--	--	--
Inerts		0	--	--	--
H_2O	1920 - 322	1598	28764	61.9	23.7
AN	950 – 11.9	938.1	87243	36.3	71.8
NB	40 – 3.8	36.2	4453	1.4	3.7
Cycl.		10	990	0.4	0.8
	Total	2582.3	121,450	100.0	100.0

This calculation ignores the solubility of nitrobenzene in the condensed aniline in the recycle gas.

Note: H_2O in the recycle gas would go through the reactor unreacted and would add to the tie H_2O in the reactor outlet. But, as the recycle gas depends on the vapour pressure (i.e. the outlet temp.) it remains as calculated.

The required flows of nitrobenzene and aniline are therefore:

Inlet Stream:

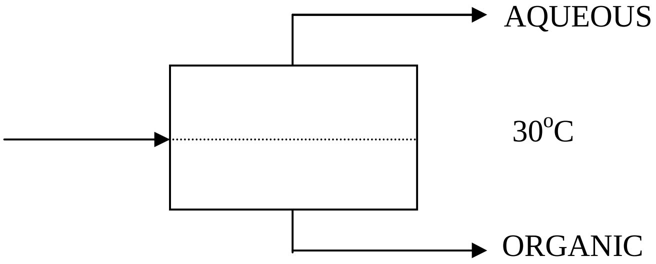
		kmol	vol %
AN		950	10.34
Cycl.		10	0.11
H_2O	1920 + 322	2242	24.42
NB		40	0.44
H_2		5640	61.42
Inerts		300	3.27
	Total	9182	100.00

An iterative calculation could be performed but it is not worthwhile.

Solution 2.7

Basis: 100 kg feed

H_20	23.8
AN	72.2
NB	3.2
Cycl	0.8
	100.0



Minor components such as nitrobenzene and aniline will be neglected in the preliminary balance.

Let the flow rate of aqueous stream be F kg per 100 kg of feed.

Flow rate of aniline and H_2O = 72.2 + 23.8 = 96.0 kg

Balance of aniline:

IN = 72.2 kg

OUT Aqueous stream $= F \text{ x } \dfrac{3.2}{100} = 0.032F$

Organic stream $= (96 - F)\left(1 - \dfrac{5.15}{100}\right) = 96 - 4.94 - F + 0.0515F$

Equating: $72.2 = 91.06 - F(1 - 0.0835)$

$F = 20.6$ kg

Organic stream = 96 – 20.6 = 75.4 kg

Nitrobenzene:

Since the partition coefficient $C_{organic}/C_{water} = 300$ more nitrobenzene leaves the decanter in the organic phase. Only a trace (≈ 3.2/300 = 0.011 kg, 11g) leaves in the aqueous phase.

Cyclo-hexylamine:

From the given solubilities, the distribution of cyclo-hexylamine is as follows:

Aqueous phase = $20.6\left(\frac{0.12}{100}\right)$ = 0.03 kg

Organic phase = $75.4\left(\frac{1}{100}\right)$ = 0.75 kg

0.78 kg (near enough)

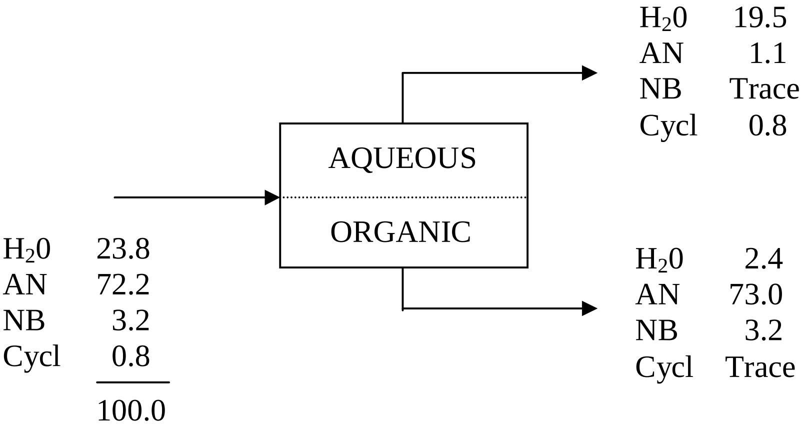
From the solubility data for aniline and water:

Aqueous phase Aniline = $20.6\left(\frac{5.15}{100}\right) = 1.1$ kg

$H_2O = 20.6 - 1.1 = 19.5$ kg

Organic phase $H_2O = 75.4\left(\frac{3.2}{100}\right) = 2.4$ kg

Aniline = 75.4 – 2.4 = 73.0 kg



H_20	23.8
AN	72.2
NB	3.2
Cycl	0.8
	100.0

H_20	19.5
AN	1.1
NB	Trace
Cycl	0.8

H_20	2.4
AN	73.0
NB	3.2
Cycl	Trace

Therefore, the H_2O and aniline flows need to be adjusted to balance. However, in this case it is probably not worth iterating.

Solution 2.8

Calculation of the feed mol fractions:

	w/w	MW	mol/100 kg h^{-1}	mol %
H_2O	2.4	18	13.3	14.1
AN	73.0	93	78.5	83.2
NB	3.2	123	2.6	2.7

Aniline in feed = 83.2 kmol h^{-1}

With 99.9 % recovery, aniline on overheads = (83.2)(0.999) = 83.12 kmol h^{-1}

Overhead composition will be near the azeotrope and so an aniline composition of 95 % is suggested.

(NB: Would need an infinitely tall column to reach the azeotrope composition)

Water composition in overheads = 100 – 95 = 5 mol %

So water carried over with the aniline = $83.12\left(\frac{5}{95}\right)$ = 4.37 kmol h^{-1}

Water leaving the column base = 14.1 – 4.37 = 9.73 kmol h^{-1}

Compositions:

		kmol h^{-1}	mol %
TOPS	AN	83.12	95.0
	H_2O	4.37	5.0
	NB	Trace	
		87.49	100.0
BOTTOMS	AN	0.08	0.64
	H_2O	9.73	77.78
	NB	2.70	21.55
		12.51	99.97

Solution 3.1

Energy $= \Delta\mathrm{P}\,\nu = \dfrac{\Delta\mathrm{P}}{\rho} = \dfrac{(100-3)}{850} \text{ x } 10^5$

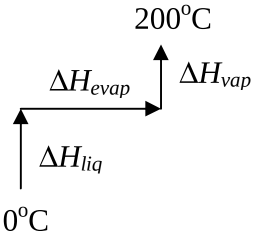
$= 11{,}412 \text{ J kg}^{-1}$

Power $= \left(\dfrac{\mathrm{J}}{\mathrm{kg}}\right)\left(\dfrac{\mathrm{kg}}{\mathrm{s}}\right)$

$= 11{,}412\left(\dfrac{1000}{3600}\right)$

$= 3170 \text{ W}$

Solution 3.2



$$\Delta H_{liq} = \int_0^{100}(4.2 - 2\,\text{x}\,10^{-3}t)dt = \left[4.2t - 2\,\text{x}\,10^{-3}\,\frac{t^2}{2}\right]_0^{100}$$

$= 420 - 10$

$= 410 \text{ kJ kg}^{-1}$

$\Delta H_{evap} = 40{,}683 \text{ J mol}^{-1}$ (From Appendix D)

$= \dfrac{40683}{18} = 2260 \text{ kJ kg}^{-1}$

From Appendix O, the specific heat of the vapour is given by:

$C_p = 32.243 + 19.238 \text{ x } 10^{-4}T + 10.555 \text{ x } 10^{-6}\,T^2 - 3.596 \text{ x } 10^{-9}\,T^3$

Where C_p is in J mol^{-1} K^{-1} and T is in K. Now 100°C = 273.15K and 200°C =373.15K.

$$\Delta H_{vap} = \int_{273.15}^{373.15}(32.243 + 19.238\,\text{x}\,10^{-4}T + 10.555\,\text{x}\,10^{-6}T^2 - 3.596\,\text{x}\,10^{-9}T^3)dT$$

$$= \left[(32.243T + 19.238\,\mathrm{x}\,10^{-4}\frac{T^2}{2} + 10.555\,\mathrm{x}\,10^{-6}\frac{T^3}{3} - 3.596\,\mathrm{x}\,10^{-9}\frac{T^4}{4}) \right]_{273.15}^{373.15}$$

= 12,330.8 – 8945.7

= 3385.1 kJ kmol^{-1}

$$= \frac{3385.1}{18}$$

= 188.1 kJ kg^{-1}

Therefore, specific enthalpy:

ΔH_{liq} = 410

ΔH_{evap} = 2260

ΔH_{vap} = 118.1

2778 kJ kg^{-1}

From Steam Tables: 2876 kJ kg^{-1}. Error = 98 kJ kg^{-1} (3.5 %).

Solution 3.3

Calculation of the enthalpy of reactions:

1.	CO	+	$\tfrac{1}{2}O_2$	$\rightarrow$	CO_2
ΔH_F (kJ mol^{-1})	-110.62		0		-393.77

ΔH_R = -393.77 – (-110.62) = -283.15 kJ mol^{-1} CO

2.	H_2	+	$\tfrac{1}{2}O_2$	$\rightarrow$	H_2O
	0		0		-242.00

ΔH_R = -242.00 – 0 = -242.00 kJ mol^{-1} H_2

3.	CH_4	+	$2O_2$	$\rightarrow$	CO_2	+	$2H_2O$
	-74.86		0		-393.77		-242.00

ΔH_R = [-393.77 + 2(-242.00)] – (-74.86) = -802.91 kJ mol^{-1} CH_4

4.	C_2H_6	+	$3\tfrac{1}{2}O_2$	$\rightarrow$	$2CO_2$	+	$3H_2O$
	-84.74		0		-393.77		-242.00

ΔH_R = [2(-393.77) + 3(-242.00)] – (-84.74) = -1428.8 kJ mol^{-1} C_2H_6

$$\begin{array}{lcccccccc} \text{5.} & C_2H_4 & + & 6O_2 & \rightarrow & 2CO_2 & + & 2H_2O \\ & 52.33 & & 0 & & -393.77 & & -242.00 \end{array}$$

$\Delta H_R = [2(-393.77) + 2(-242.00)] - 52.33 = -1323.87$ kJ mol^{-1} C_2H_4

$$\begin{array}{lcccccccc} \text{6.} & C_6H_6 & + & 7\tfrac{1}{2}O_2 & \rightarrow & 6CO_2 & + & 3H_2O \\ & 82.98 & & 0 & & -393.77 & & -242.00 \end{array}$$

$\Delta H_R = [6(-393.77) + 3(-242.00)] - 82.98 = -3171.6$ kJ mol^{-1} C_6H_6

Composition (mol %):

CO_2: 4, CO: 15, H_2: 50, CH_4: 12, C_2H_6: 2, C_2H_4: 4, C_6H_6: 2, N_2: 11.

Basis: 100 mol

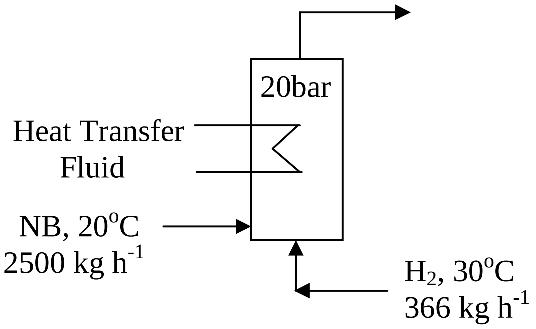
Component	Quantity	$-(\Delta H_R)$	H (kJ)	
CO_2	4	--		
CO	15	283.15	4247.25	
H_2	50	242.00	12100.00	
CH_4	12	802.91	9634.92	
C_2H_6	2	1428.8	2857.60	
C_2H_4	4	1323.87	5295.48	
C_6H_6	2	3171.60	6343.20	
N_2	11	--		
	100		40478.45	(kJ/100 mol)

Therefore, $H = (40478.45)(10) = 404784.5$ kJ kmol^{-1}

Gross CV (kJ m^{-3}) = $\dfrac{404{,}784.5}{22.4} = 18{,}071$ kJ m^{-3} (= 485 BTU ft^{-3})

To calculate the Net CV, subtract the heat of vapourisation of the H_2O burned.

Solution 3.4



Molecular weight of nitrobenzene = 123 and H_2 = 2

Molar flow of nitrobenzene $= \dfrac{2500}{(123)(3600)} = 5.646 \times 10^{-3}$ kmol s^{-1}

Molar flow of $H_2 = \dfrac{366}{(2)(3600)} = 50.833 \times 10^{-3}$ kmol s^{-1}

Partial pressure of nitrobenzene $= \left(\dfrac{5.646 \times 10^{-3}}{[5.646 \times 10^{-3} + 50.83 \times 10^{-3}]}\right) 20 = 2.0$ bar

Using the Antoine Equation: $\ln P = A - \dfrac{B}{T + C}$

The Antoine constants are obtained from Appendix D. (2 bar = 1500 mm Hg)

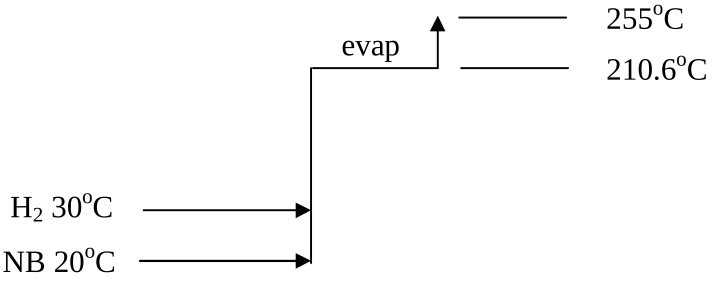
$$\ln(1500) = 16.1484 - \frac{4032.6}{T - 71.81}$$

$$7.313 - 16.1484 = \frac{-4032.6}{T - 71.81}$$

$$T - 71.81 = \frac{-4032.6}{-8.8352} = +456.4$$

T = 528 K = 255°C

The boiling point of nitrobenzene at 1 atm = 210.6°C (Appendix D)



The specific heat capacity of the nitrobenzene liquid can be estimated using Chueh and Swanson’s method.

$-\overset{\vert}{\underset{\vert}{\mathrm{CH}}}$	$-\overset{\vert}{\mathrm{C}}$	$\mathrm{N}<$	$-\mathrm{O}-$	
(18.42 x 5)	12.14	18.84	(35.17 x 2)	Total = 193 kJ $kmol^{-1}$ C^{-1}

The specific heat capacity of the nitrobenzene gas:

	a	*b* x 10^2	*c* x 10^4	*d* x 10^6
$\mathrm{HC}<$	-6.1010	8.0165	-0.5162	0.01250
(x 5)	-30.505	40.083	-2.581	0.0625
$-\mathrm{C}-$	-5.8125	6.3468	-0.4776	0.01113
NO_2	4.5638	11.0536	-0.7834	0.01989
	-31.7537	57.4829	-3.8420	0.0935

Nitrobenzene:

H_{liq} = (5.646 x 10^{-3})(193)(210.6 – 20) = 208 kW

$$\Delta H_{gas} = 0.005646 \int_{484}^{528} (-31.7537 + 57.4829\times10^{-2} T - 3.842\times10^{-4} T^2 + 0.0935\times10^{-6} T^3)\, dT$$

= 43 kW

$$\Delta H_{evap} = \left(44{,}031 \frac{\mathrm{kJ}}{\mathrm{kmol}}\right)\left(5.636\times10^{-3} \frac{\mathrm{kmol}}{\mathrm{s}}\right) = 248.15 \text{ kW}$$

H_2:

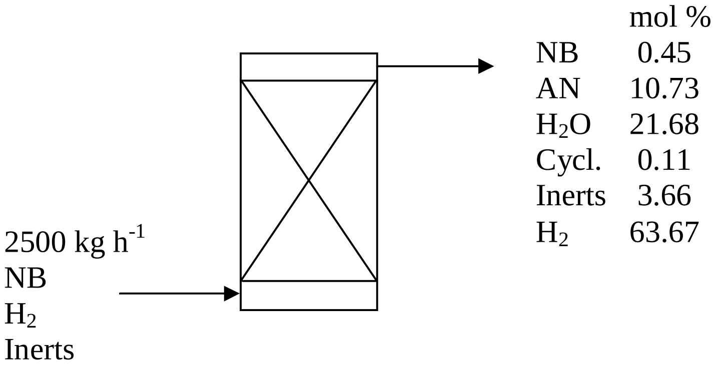
$$\Delta H_{gas} = 0.05083 \int_{303}^{528} (27.143 + 92.783\times10^{-3} T - 1.38\times10^{-5} T^2 + 76.45\times10^{-10} T^3)\, dT$$

= 730 kW

Therefore: Total ΔH = 208 + 43 + 248 + 730 = 1229 kW

Note: It is not worth correcting the heat capacities for pressure.

Solution 3.5



Nitrogen Balance:

Molar flow of nitrobenzene = $\dfrac{2500}{(123)(3600)}$ = 5.646×10^{-3} kmol s^{-1}

Therefore, katoms N = 5.646×10^{-3} s^{-1}

Let the total mass out be x, then:

$5.646 \times 10^{-3} = x\left[\dfrac{0.45+10.73+0.11}{100}\right]$

$x = 0.050$ kmol s^{-1}

		H_2 reacted
Aniline produced	$= (0.05)\left(\dfrac{10.73}{100}\right) = 0.00536$	0.0161
Cyclo-hexylamine produced	$= (0.05)\left(\dfrac{0.11}{100}\right) = 0.000055$	0.0003
		0.0164 kmol s^{-1}
Unreacted H_2	$= (0.05)\left(\dfrac{63.67}{100}\right)$	0.0318
So, total H_2 In =		0.0482 kmol s^{-1}

Now, $\Delta H_{reaction}$ = 552,000 kJ $kmol^{-1}$ (Appendix G8)

From ΔH_f (Appendix D)

NB	-67.49 kJ mol^{-1}
AN	86.92
H_2O	-242.00

$\Delta H_{reaction}$ = Σ products – Σ reactants

= [86.92 + 2(-242.00)] – (-67.49)

= -329.59 kJ mol^{-1}

= 329,590 kJ kmol^{-1}

Reactions: $C_6H_5NO_2 + 3H_2 \rightarrow C_6H_5NH_2 + 2H_2O$

$C_6H_5NH_2 + 3H_2 \rightarrow C_6H_{11}NH_2$

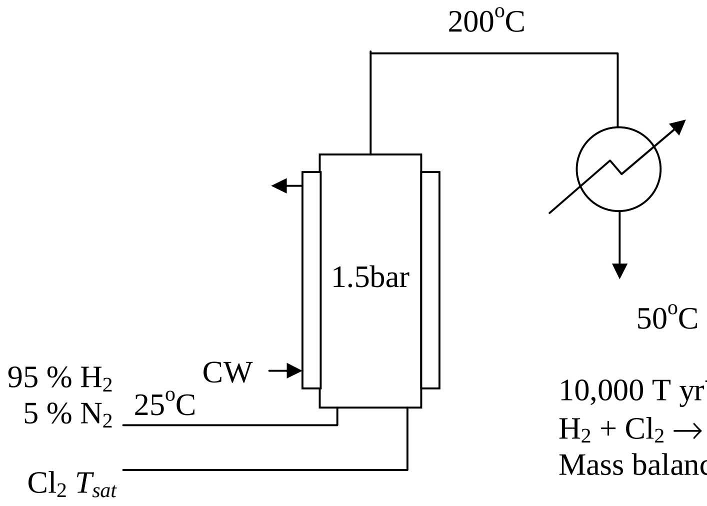
The second reaction can be ignored since it represents a small fraction of the total.

The problem can be solved using the ENRGYBAL program. Heat capacities can be found in Appendix D and calculated values for nitrobenzene obtained from Solution 3.4.

Solution 3.6

A straight-forward energy balance problem. Best to use the energy balance programs: ENERGY 1, page 92 or ENRGYBAL, Appendix I, to avoid tedious calculations. Data on specific heats and heats of reaction can be found in Appendix D.

What follows is an outline solution to this problem.



10,000 T yr^{-1} HCl
$H_2 + Cl_2 \rightarrow 2HCl$
Mass balance (1 % excess) gives feed.

Solution:

1. T_{sat} for Cl_2 from Antoine Equation (Appendix D),
2. $\Delta H_{reaction}$ from the HCl heat of formation,
3. C_p's from Appendix D,

4. Reactor balance to 200°C (4 % free Cl_2),

5. Datum temperature 25°C,

6. Ignore pressure effects on C_p's.

Reactor:

IN - 1. $H_2 + N_2$ = zero (at datum temperature),

2. Cl_2 at T_{sat} (note as gas, $\Delta H_{reaction}$ for gases),

3. $\Delta H_{reaction}$ at 25°C (96 % Cl_2 reacted).

OUT - 1. $HCl + Cl_2 + H_2$ (excess) + N_2 at 200°C,

2. Cooling in jacket.

Cooler:

IN - 1. Reactor outlet H,

2. 4 % Cl_2 reacted ($\Delta H_{reaction}$).

OUT - 1. Sensible heat of HCl, H_2 (excess) and N_2,

2. Heat to cooling water.

Check on T_{sat}:

$$\ln(1.5 \times 750) = 15.9610 - \frac{1978.32}{T_{sat} - 27.01}$$

$$\frac{1978.32}{T_{sat} - 27.01} = 15.9610 - 7.026$$

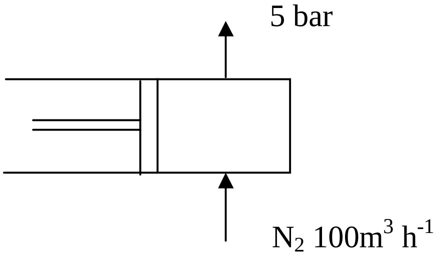
$$T_{sat} = \frac{1978.32}{8.935} + 27.01 = 248.4 \text{ K}$$

$T_{sat} = -24.6^{o}$C (Within the temperature limits)

The Cl_2 may need preheating.

Solution 3.7

As $P_2 < P_{critical}$, the simplified equation can be used.



γ = 1.4 for air.

$$-w = P_1 v_1 \left(\frac{n}{n-1}\right)\left\{\left(\frac{P_2}{P_1}\right)^{\frac{n-1}{n}} - 1\right\}$$

where: $n = \dfrac{1}{1-m}$ and $m = \dfrac{\gamma - 1}{\gamma E_p}$.

Compression ratio = 10 – from Figure 3.7, E_p = 86 %.

$$T_2 = T_1 \left(\frac{P_2}{P_1}\right)^m$$

$$m = \frac{1.4-1}{(1.4)(0.86)} = 0.33,\ n = \frac{1}{1-0.33} = 1.49\,.$$

$$T_2 = (20+273)\left(\frac{10}{1}\right)^{0.33} = 626 \text{ K} = 353^{\circ}\text{C}$$

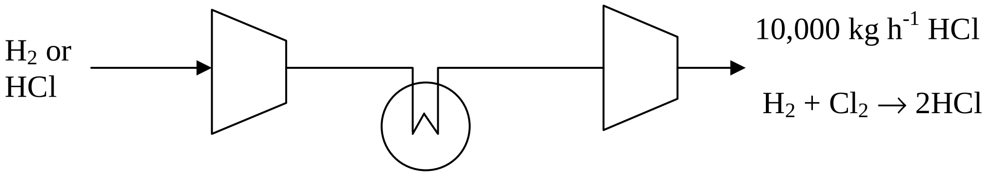
In practice the compressor cylinder would be fitted with a cooling jacket.

v_1 = 100 m^3 h^{-1} = 0.0278 m^3 s^{-1}

$$-w = (10^5)(0.0278)\left(\frac{1.49}{1.49-1}\right)\left\{(10)^{\frac{1.49-1}{1.49}} - 1\right\}$$

$$= 9.6 \text{ kW} \quad (\text{Say } 10 \text{ kW})$$

Solution 3.8



Burner operating pressure, 600 kN m^{-2} required. Take burner as operating at 1 atm. = 10^2 kN m^{-2} g or 600 kN m^{-2} g.

H_2 is compressed from 120 kN m^{-2} to 600 kN m^{-2}.

Pressure ratio = $\frac{600}{120} = 5$

Intermediate pressure = $\sqrt{P_1P_2} = \sqrt{(120)(600)} = 268$ kN m^{-2}

Note: For H_2 the inlet temperature will not be the same as the intercooler outlet so the cool stage should be calculated separately.

A material balance gives the H_2 flow. The 1 % excess H_2 is ignored in the HCl compressor calculation.

Material balance:

HCl produced = $\frac{10{,}000}{(36.5)(3600)}$ = 0.0761 kmol s^{-1}

H_2 required = $\left(\frac{0.0761}{2}\right)1.01$ = 0.0384 kmol s^{-1}

Cl_2 required = $\left(\frac{0.0761}{2}\right)$ = 0.0381 kmol s^{-1}

Excess H_2 = 0.0384 – 0.0381 = 0.0003 kmol s^{-1}

The simplified equations (3.36a and 3.38a) can be used since conditions are far removed from critical.

Take $\gamma = 1.4$ since both H_2 and HCl are diatomic gases.

$$m = \frac{(1.4-1)}{(1.4)(0.7)} = 0.408 \qquad (3.36a)$$

$$n = \frac{1}{1-0.408} = 1.689 \qquad (3.38a)$$

$$-w = P_1v_1\left(\frac{n}{n-1}\right)\left\{\left(\frac{P_2}{P_1}\right)^{\frac{n-1}{n}} - 1\right\} \qquad (3.31)$$

H_2:

1st Stage: $$v_1 = \left(\frac{2}{22.4}\right)\left(\frac{1.013\times10^5}{120\times10^3}\right)\left(\frac{298}{273}\right) = 0.0823 \text{ m}^3 \text{ kg}^{-1}$$

$$-w_1 = (120\times10^3)(0.0823)\left(\frac{1.689}{1.689-1}\right)\left\{\left(\frac{268}{120}\right)^{\frac{1.689-1}{1.689}} - 1\right\} = 9{,}391 \text{ J kg}^{-1}$$

2nd Stage: $$v_1' = \left(\frac{2}{22.4}\right)\left(\frac{1.013\times10^5}{268\times10^3}\right)\left(\frac{323}{273}\right) = 0.0399 \text{ m}^3 \text{ kg}^{-1}$$

$$-w_2 = (268\times10^3)(0.0399)\left(\frac{1.689}{1.689-1}\right)\left\{\left(\frac{600}{268}\right)^{\frac{1.689-1}{1.689}} - 1\right\} = 10{,}204 \text{ J kg}^{-1}$$

Power = (9,391 + 10,204)(0.0384)(2) = 1505 W = 1.505 kW

HCl:

Take both stages as performing equal work with the same inlet temperature.

$$P_i = \sqrt{P_1 P_2} = \sqrt{(101.3)(600)} = 246.5 \text{ kN m}^{-2} \qquad (3.39)$$

$$v_1 = \left(\frac{36.5}{22.4}\right)\left(\frac{1.013\times10^5}{1.013\times10^5}\right)\left(\frac{323}{273}\right) = 1.927 \text{ m}^3 \text{ kg}^{-1}$$

$$-w_1 = (1.013\times10^5)(1.927)\left(\frac{1.689}{1.689-1}\right)\left\{\left(\frac{600}{101.3}\right)^{\frac{1.689-1}{1.689}} - 1\right\} = 510{,}173 \text{ J kg}^{-1}$$

Power = (510,173)(0.0761)(36.5) = 1,417,082 W = 1417 kW

It is necessary to divide by the efficiency to get the actual power but it is clear that the best choice is to compress the H_2 and operate the burner under pressure.

Check:

Temperature of saturated Cl_2 at 600 kN m^{-2}.

$$\ln\left(\frac{600\times10^3}{133.32}\right) = 15.9610 - \frac{1978.32}{T - 27.01}$$

$$\frac{1978.32}{T-27.01} = 15.9610 - 8.412$$

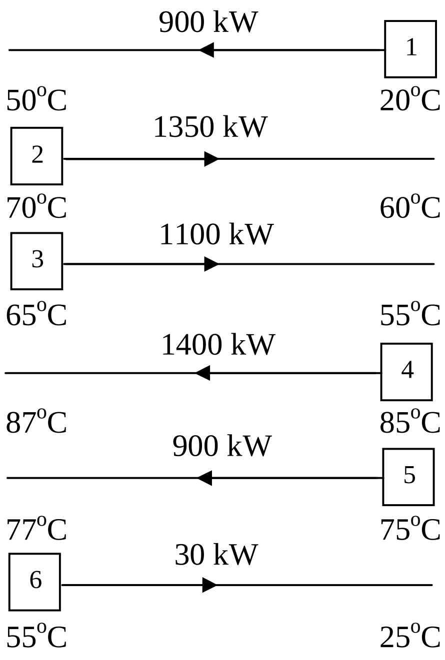
$T = 262 + 27.01 = 289\ \text{K} = 16^{o}\text{C}$

Solutions 3.9 and 3.10 .

Refer to example 3.17 and the worked solution to problem 3.12

Solution 3.11

Streams:		T_s (^{o}C)	T_t (^{o}C)	C_p (kW ^{o}C^{-1})
Preheater	C	20	50	30
Condenser 1	H	70	60	135
Condenser 2	H	65	55	110
Reboiler 1	C	85	87	700
Reboiler 2	C	75	77	450
Cooler	H	55	25	1



For $T_{min} = 10^{o}$C

$T_{int} = T_{out} + 5$ (cold)

$T_{int} = T_{out} - 5$ (hot)

Stream	Type	T_{act}		T_{int}	
1	C	20	50	25	55
2	H	70	60	65	55

3	H	65	55	60	50
4	C	85	87	90	92
5	C	75	77	80	82
6	H	55	25	50	20

Ranked Streams

(^{o}C)		kW	Cascade	Add
92		0		2300
90	C 4	+1400	-1400	900
82		0	-1400	900
80	C 5	+900	-2300	0
65		0	-2300	0
60	H 2	-550	-1750	550
55	H 3	-1225	-525	1775
50	C 1	-400	-125	2175
25		725	-850	1450
20	H 6	5	-855	1445

$$\Delta H = \left(\sum \left(C_p\right)_C - \sum \left(C_p\right)_H\right)\Delta T$$

Hot Utilities = 2300 kW

Cold Utilities = 1445 kW

Pinch = 60 – 82^{o}C

Solution 4.1

Basis 100 kmol benzene at reactor inlet

Reactor:

Cl_2 at reactor inlet = (100)(0.9) = 90 kmol
C_6H_6 converted = (100)(0.553) = 55.3 kmol
C_6H_5Cl produced = (55.3)(0.736) = 40.70 kmol
$C_6H_4Cl_2$ produced = (55.3)(0.273) = 15.10 kmol

Cl_2 reacted = 40.70 + 2(15.10) = 70.90 kmol
HCl produced = 70.90 kmol
Cl_2 unreacted = 90 - 70.90 = 19.10 kmol

Separator:

Gas phase:	Cl_2	19.10 kmol
	HCl	70.90
Liquid phase:	C_6H_6 = 100 – 55.3	44.70 kmol
	C_6H_5Cl	40.70
	$C_6H_4Cl_2$	15.10

Absorber:

HCl In = (70.90)(36.5) = 2588 kg

Water for 30% w/w acid $= \dfrac{2588}{0.30}$ = 8626 kg

Therefore, Solution Out = 11,214 kg

Neglect water vapour carried over with chlorine

Assume all HCl absorbed together, with 2 percent of the chlorine

Cl_2 recycled = (19.10)(0.98) = 18.72 kmol

Distillation:

Feed:	C_6H_6	44.70 kmol
	C_6H_5Cl	40.70
	$C_6H_4Cl_2$	15.10

Overheads:

With 0.95% recovery, $C_6H_6 = (44.70)(0.95) = 42.47$ kmol

Bottoms:	C_6H_6	$= 44.70 - 42.47$	2.33 kmol
	C_6H_5Cl		40.70
	$C_6H_4Cl_2$		15.10

Reactor with recycle feeds –

Fresh feeds:	C_6H_6	$= 100 - 42.47$	57.53 kmol
	HCl	$= 90 - 18.72$	71.18 kmol

Scaling factor –

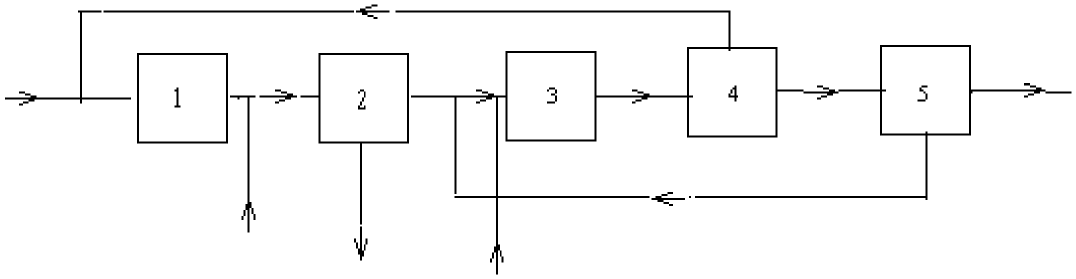
Product required $= 100 \text{ t d}^{-1} = \dfrac{1000}{24} = 41.67 \text{ kg h}^{-1} = \dfrac{41.67}{112.5} = 0.37 \text{ kmol h}^{-1}$

So, 57.53 kmol fresh feed of benzene to the reactor produces 40.70 kmol of product.

Therefore, scaling factor for flow sheet $= \dfrac{0.37}{40.70} = 0.0091$

I would use a slightly higher factor to give a factor of safety for losses, say 0.0095.
A second, and possibly a third, column would be need to separate the monochlorobenzene from the dichlorobenzene and unreacted benzene – see Chapter 11, Section 11.6.2.

Solution 4.2



1. Reactor
2. MTBE column
3. Absorber
4. MeOH distillation
5. Recycle splitter (tee)

g10k = feed stock + MeOH
g20k = pseudo feed MTBE
g30k = water make-up

Components (k’s):
1. C4’s, other than isobutane
2. methanol (MeOH)
3. isobutane
4. MTBE
5. water

Number of split fraction coefficients = (N- 1) + R = (5 – 1) + 2 = 6

Equations (matrix) 5 units

$$\begin{array}{ccccccc} & 1 & 2 & 3 & 4 & 5 & g\text{'s} \\ 1 & 1 & \alpha_{21k} & 0 & 0 & 0 & g_{10k} \\ 2 & 0 & 1 & \alpha_{32k} & 0 & 0 & g_{20k} \\ 3 & 0 & 0 & 1 & \alpha_{43k} & 0 & g_{30k} \\ 4 & \alpha_{14k} & 0 & 0 & 1 & \alpha_{54k} & \\ 5 & 0 & 0 & \alpha_{35k} & 0 & 1 & \end{array}$$

Estimation of α's and g's –

Basis 100 kmol h^{-1} feed-stock

Spilt fraction coefficients, α 's, subscripts give without punctuation.

k = 1: C4's, other than isobutane.

Assume they pass through unchanged, no reaction and no absorption.

211 = 1.0

431 = 0.0 (sent to storage, other uses)

321 = 0.0

541 = 0.0

351 = 0.0

141 = 0.0

Fresh Feeds, g101 = ΣC4's = 2 + 31 + 18 = 51 kmol

k = 2: MeOH

With 10% excess and 97% conversion,

Feed of iC4		= 49 kmol
So, Inlet MeOH	= (1.1)(49)	= 53.9 kmol
MeOH reacted	= (0.97)(49)	= 47.5 kmol
MeOH Out	= 53.9 – 47.5	= 6.4 kmol.

212 = $\frac{6.4}{53.9}$ = 0.12
322 = 1.0 MTBE (pure so, negligible loss of MeOH)
432 = 0.99 (99% recovery)
542 = 0.01 (99% recovery)
142 = 0.99 (99% recovery)
352 = 0.9 (10% purge)

Fresh feed, g102 = 49 (put equal to isobutane in feed and adjust after first run to allow for losses)

k = 3: isobutene
213 = 1 – 0.97 = 0.03 (97% conversion)
323 = 1 – 0.99 = 0.01 (99% recovery)
433 = 1.0
543 = 1.0 (no MTBE)
143 = 1.0
153 = 1.0

Fresh feed, g103 = 49 kmol

k = 4: MTBE
214 = 1.0
324 = 0.005 (99.5% recovery in column)
434 = 0.0 (assumed not absorbed)
544 = 1.0
354 = 1.0
144 = 1.0

Fresh feed, 104 = 47.5 kmol, (produced in reactor)

k = 5: water
215 = 1.0
325 = 1.0
435 = 0.965 (allow for carry over with C4’s, see note 1.)
545 = 0.99 (99% recovery)
355 = 0.9 (10% purge)
145 = 0.01 (recycle?)

Fresh feed, 302 = 8 kmol (see note 2)

Notes:

Carry over of water with C4’s from column.

Vapour pressure of water at 30^{o}C = 0.0424 bar (approximately 4.2%)

C4’s flow = 51 kmol

Loss of water $= \frac{51}{1-0.042}$ = 52.24 kmol

Water flow rate, recycle, = 64 kmol (notes)

Split fraction $= \frac{2.24}{64}$ = 0.035

Water Fresh Feed:

Concentration of MeOH at absorber base = 10%

MeOH = (0.13)(49) = 6.37 kmol

Total flow $= \frac{6.37}{0.1}$ = 63.7 kmol

10 % purge = (63.7)(0.1) = 6.4 kmol

Water = (0.9)(6.4) = 5.8 kmol

If we add the loss with C4’s leaving column, Total = 5.8 + 2.2 = 8 kmol.

Solution

Use spread sheet or the program MASBAL to solve. My solution, using the split fractions and fresh feeds given above, is set out below. The table shows in flows at the inlet of each unit, rounded to one place, (in kmol h^{-1}).

Component	Unit 1	Unit 2	Unit 3	Unit 4	Unit 5
1. C’4s	51	51	51	0	0
MeOH	55.9	6.7	6.8	6.7	0.1
iC4	49.0	1.5	trace	trace	trace
MTBE	0	47.5	0.2	0	0
H2O	0.6	0.6	57.6	55.6	54.6
Total	156.5	107.2	115.6	62.3	54.6

The other stream flows can be obtained form mass balances round the units or by including dummy unit in the information diagram

Iterate on split fraction and fresh feeds, as necessary to match the constraints.
For example, the water purge seems low.

Solution 4.3
What follows is a partial solution and notes.

Careful choice of the starting point will avoid the need for iteration.
Start at the inlet to the decanter, where the composition is fixed at the ternary azeotrope.
Take the basis as 100 kmol h^{-1} feed to the decanter. Let F1 be the flowrate of decanter stream returned to the first column and F2 the stream going to the second column. A component material balance will determine these stream flows.

Benzene $54 = (F1)(0.74) + (F2)(0.04)$
Water $22 = (F1)(0.04) + (F2)(0.61)$
Solving gives: F1 = 71.3 kmol h^{-1} and F2 = 28.7 kmol h^{-1}

All the benzene going to Column 1 from the decanter leaves in the column overhead and so the overhead rate, F3 = $71.3\left(\frac{0.74}{0.54}\right)$ = 97.7 kmol h^{-1}

The balance to make up the 100 kmol h^{-1} to the decanter is the overheads from Column 2.

F8 = 100 – 97.7 = 2.3 kmol h^{-1}

No water leaves the base of Column 1 and so the water entering the column in the feed, F5, and the stream from the decanter, go overhead.

A water balance gives F5: 0.11 F5 + (71.3)(0.04) = (97.7)(0.22)

F5 = 68.5 kmol h^{-1}

A balance on ethanol gives the bottoms flow, F6:

(68.5)(0.89) + (71.3)(0.22) = F6 + (97.7)(0.24)

F6 = 53.2 kmol h^{-1}

The only source of this product ethanol is the fresh feed to the column, F7 and so:

$$F7 = \frac{53.2}{0.89} = 59.8 \text{ kmol h}^{-1}$$

So the recycled overhead product from the third column, F4 is:

F4 = 68.5 – 59.8 = 8.7 kmol h^{-1}

All the water leaves the system in the bottoms for Column 3 and so the bottoms from this column, F8, will be:

F8 = (59.8)(0.11) = 6.6 kmol h^{-1}

The flow sheet is to be drawn for a production rate of 100 kmol/h of absolute alcohol, so the scaling factor required is $\frac{100}{53.2}$ = 1.88. (Say 1.9)

The make up benzene can be added in the stream from the decanter to Column 1.

Solution 4.4

Notes/Hints:

There are three main pieces of equipment involved in the flow sheet calculations: the reactor, absorber, and stripper, and two minor pieces: the vent scrubber and dryer.

The reactor flows can be calculated from the stoichiometry of the reaction.

It is not necessary to make repetitive calculations to determine the flow of recycled acid to the absorber. The recycle flow is fixed by the change in the specified acid concentration from inlet to outlet.

In the dryer, the purge stream rate is determined by the amount of water removed and the acid concentration. The acid recycle rate will be a design variable in the design of the drying column.

Solution 4.5

Refer to the solution to Problem 4.2

Solution 4.6

Refer to the solution to Problem 4.2

Solution 5.1

See section 5.3 for guidance. Where flow control is not required, any type giving a positive closure could be used: plug, gate or ball. The final selection would depend on the valve size, materials and cost.

Example: The block valves could be plug or ball. The valve on the by-pass stream would need to be a globe valve to give sensitive flow control.

Solution 5.2

See Example 5.2 and the solution to Problem 5.4

Solution 5.3

See Example 5.2 and the solution to Problem 5.4. Remember that power is taken from a turbine, so the work term should be positive.

Solution 5.4

Equivalent length of pipe, use values from table 5.3

Inlet line		Outlet line	
inlet	= 25	outlet	.= 50
elbows 6 x 40	= 240	10 x 40	= 400
gate valves, open	= 7.5	4 x 7.5	= 30
Total =	272.5		480

$L' = (25 + 250) + (272.5 + 480) \times 75 \times 10^{-3} = 231.4$ m

Static pressure

$(z_1 - z_2) = (4 - 6) = -2$ m fluid

$(P_1 - P_2) = (1.05 - 1.3) = -0.25$ bar $= (0.25 \times 10^5)/(9.8 \times 875) = -2.92$ m fluid

Total $-2 + (-2.92) + -4.92$

Take flow-rate, initially, as 36.3 m^3/h

cross-sectional area $\Pi/4\ (75 \times 10^{-3}) = 4.42 \times 10^{-3}\ m^2$

vel, $u = 36.3/3600 \times 1/4.42 \times 10^{-3} = 2.28$ m/s

relative roughness, $e/d = 0.046/75 = 0.006$

$Re = \underline{875 \times 2.28 \times 75 \times 10^{-3}} = 102{,}483 = 1.02 \times 10^5$ (5.4)

from Fig 5.7, $f = 0.0025$

$\Delta P_f = 8 \times 0.0025(331.4/75\times10^{-3})875 \times 9.28^2/2 = 200{,}987\ N/m^2$

as liquid head = 200987/(9.8 x 875) = 24.14 m

drop across control valve = 35/(9.8 x 875) = 4.08 m

Total static head = 4.08 + 4.92 = 9.0 m

Total head at this flow rate = 9 + 24.14 = 33.1. m

Repeat calculation for various flow rates

Flow m^3/h	0	27.3	36.3	45.4
Static pressure	9	9	9	9
Dynamic press drop	0	13.6	24.1	37.7
Total	9	22.6	33.1	46.7

Plotting this operating curve on the pump characteristic gives the operating point as 29.5 m at 33.0 m^3/h

Suction head

$H = 2\ m$, $P = 1.05 \times 10^5\ N/m^2$

$L' = 25 + (275.5 \times 75 \times 10^{-3}) = 45.7\ m$

$u = 33/3600 \times 1/ 4.42\times10^{-3} = 2.07\ m/s$

$\Delta P_f = 8 \times 0.0025 (45.7/75 \times 10^{-3}) 875 \times 2.07^2/2 = 22{,}846\ N/m^2$

$P_v = 25 \times 10^3\ N/m^2$

$NPSH = 4 + 1.05 \times 10^5/ (875 \times 9.8) - 22846/(875 \times 9.8) - 25 \times 10^3/(875 \times 9.8)$

$= 10.7\ m$ (OK)

Solution 5.5

Close control of the reactor temperature is important. If control is lost the reactor seals could be blown and carcinogenic compounds released into the atmosphere. Interlocks and alarms should be included in the control scheme.

Solution 5.6

<u>Notes on a possible control scheme.</u>

1. The feed is from storage, so a flow controller should be installed to main constant flow to the column. A recorder could be included to give a record of the quantity of feed processed.

2. A level controller will be needed to main a liquid level in the base of the column and provide the NPSH to the pump. The level could be controlled by regulating the bottoms take-off with a valve, situated on the pump discharge, or by controlling the live steam flow to the column. Temperature control of the steam supply would not be effective, as there would be virtually no change in temperature with composition at the base. The effluent is essentially pure water

3. A level controller would be needed to maintain a level in the condenser, or separating vessel, if one were used. The level would be controlled with a valve in the product take-off line.

4. The primary control of quality would be achieved by controlling the reflux rate to meet the product purity specified. Temperature control could be used but the sensing point would need to be sited at a point in the column where there is a significant change in temperature with composition. A better arrangement would be to use a reliable instrument, such as a chromatography, to monitor and control composition. A recorder could be included to give a record of the product quality.

5. As acetone is easily separated from water, it should not be necessary to control the bottom composition directly. Any effluent above the specification that slipped through would be blended out in the effluent pond.

6. A pressure controller would be needed on the vent from the condenser, to maintain the column pressure

Solution 6.1

See Example 6.2

Solution 6.2

Use the step counting methods given in Section 6.5.2

Gas phase reaction but liquid separation and purification and so try equation 6.3:

From Appendix G8, conversion is around 98%.

$N = 6$

$Q = 20{,}000 \text{ t y}^{-1}$

$s = 0.98$

$C = (130{,}000)(6)\left(\dfrac{20{,}000}{0.98}\right)^{0.3}$ = £15,312,136 (Say 15 million pounds)

Try the equation for gas handling processes, equation 6.5:

$C = (13{,}000)(6)(20{,}000)^{0.615}$ = £34,453,080 (Say 35 million pounds)

So, the true cost is probably around 25 million pounds

Solution 6.3

See Example 6.1

Solution 6.4

See Example 6.1

Solution 6.5

See equipment cost estimates in Example 6.4

Solution 6.6

See estimates of heat exchanger costs in Example 6.4

Solution 6.7

Follow the procedure used to estimate the equipment costs in Example 6.4

Solution 6.8

Capital cost:

Tower (0.5m dia., 4m high, CS). From Fig 6. = £4000

Packing, volume $= \frac{\pi}{4}(3)0.5^2$ $= 0.59\ \text{m}^3$

Cost, Table 6.3, for 25mm ceramic $= 840\ £\ \text{m}^{-3}$

Cost of packing = (0.59)(840) = £496

Storage tank, cost as process tank, 5m^3, CS.

Table 6.2, Cost $= (1450)(5^{0.6})$ = £3800

Total = £8296 (Say £8500)

Take a Lang factor of 4.7 for fluids processing.

Therefore, Total Fixed Capital = (4.7)(8500) = £39,950 (Say £40,000)

Working capital would be negligible.

Operating costs:

From Table 6.6, Raw materials (solvent) $= (10)(365)\left(\frac{20}{100}\right)$ = £730 y^{-1}

Other variable costs negligible

Fixed costs:

Capital charges $= 40{,}000\left(\frac{15}{100}\right)$ = £6000

Assume no extra labour or supervision needed, and laboratory costs negligible.

Total operating costs = 730 + 6000 = £6730

Round to £7000 to cover factors neglected.

Savings:

Product loss = (0.7)(24)(365) = 6132 kg y^{-1}

Recovery $= 6132\left(\frac{80}{100}\right)$ = 4906 kg y^{-1}

Savings = (4906)(5) = £24,528 y^{-1}

‘Profit’ = Savings – Operating Cost = 24528 – 7000 = £17528 y^{-1}

Return on investment $= \dfrac{17528}{40000}$ = 44%

Pay-back time $= \dfrac{40000}{17528}$ = 2.3 years

So the project would be well worthwhile considering. However, as the annual savings are not large and so unless the product was in short supply, it may not be worth the time and trouble.

Solution 6.9

Estimate the cost of a suitable packaged boiler from Table 6.2. Assume the price includes any water treatment facilities needed.

Estimate the used per tonne of steam produced. Take the cost of fuel from Table 6.5.

Use the procedure set out in Table 6.6 to estimate the production cost. Main cost will be fuel (raw material), manpower and maintenance. Other costs can be neglected.

Solution 6.10

See Example 6.4

Solution 6.11

Refer to Example 6.6.

Year	Capital	Income	Net Cash Flow
0			
1	-1500000	0	-1500000
2	-1500000	0	-1500000
3		700000	700000
4		700000	700000
5		700000	700000
6		700000	700000
7		700000	700000
8		700000	700000
9		700000	700000
10		700000	700000
11		700000	700000
12		700000	700000

Cumulative NPV = £1,352,072

The NPV formula from MS WORKS was used to calculate the NPV. Most spreadsheets include such procedures.

To find the DCFRR vary the discount rate until the cumulative NPV = zero.

Here, DCFRR = 17%.

Solution 7.1

See Section 7.4.1

Solution 7.2

1. Carbon steel, schedule 40, life 3 years

   Number of replacements = 3

   Cost = 3 x L(3 + 10) = £39L (where L = pipe length)

2. Carbon steel, schedule 80

   Wall thickness = 0.5(114.5 – 97.2) = 8.6 mm

   Schedule 40 wall thickness = 6.0mm

   So, schedule 80 should last at least (3)(8.6/6.0) = 4.3 years and so only needs replacing twice. Cost = 2 x L(5 + 10) = £45L

3. Stainless steel, replace once

   Cost = 1 x L(15 + 10) = 25L

Therefore, the best choice is stainless steel.

Solutions 7.3, 7.4, 7.5, 7.6 and 7.7

Select suitable materials using the corrosion chart in Appendix C.

Then check the properties of the materials selected in Section 7.8, or other suitable references.

Solution 7.8

See Section 7.7.

Stainless steel, type 304 would be suitable. Aluminium could be considered as an alternative.

Solution 8.1

Refer to Section 8.6.1

2-butanol: C_4H_9OH Molecular mass = 48 + 10 + 16 = 74

Now $\rho_b = \frac{M}{V_m}$ (8.1)

V_m from Table 8.6:	C = (0.0148)(4)	= 0.0592
	H = (0.0037)(10)	= 0.0370
	O = 0.0074	= 0.0074
	Total	= 0.1036

Therefore, $\rho_b = \frac{74}{0.1036}$ = 714 kg m^{-3}

Solution 8.2

Refer to Section 8.6.2

Water at 100 bar and 550°C

From Appendix D: T_c = 647.3 K and P_c = 220.5 bar

So, $T_r = \frac{500+273}{647.3} = 1.19$

$P_c = \frac{100}{220.5} = 0.454$

From Fig. 3.8, z = 0.9

$PV = znRT$ (8.5)

P = 100 bar = 10,000 kN m^{-2}

R = 8.314 kJ $kmol^{-1}$ K^{-1}

$$V = \frac{(0.9)(1)(8.314)(773)}{10^4} = 0.578 \text{ m}^3 \text{ kmol}^{-1}$$

$$= \frac{0.578}{18} = 0.0321 \text{ m}^3 \text{ kg}^{-1}$$

$$\rho = \frac{1}{V} = \frac{1}{0.0321} = 31.1 \text{ kg m}^{-3}$$

Steam tables give the specific volume at these conditions as 0.0328 m^3 kg^{-1}

Solution 8.3

See Section 8.7.1

Solution 8.4

n-butane: C_4H_{10} Molecular mass = 48 + 10 = 58

Liquid:

Estimate C_p using the group contributions given in Table 8.3

$-CH_3$ (36.84)(2) = 73.68

$-CH_2-$ (30.40)(2) = 60.80

Total = 134.48 kJ $kmol^{-1}$ °C^{-1}

$$= \frac{134.48}{58} \qquad = 2.32 \text{ kJ kg}^{-1} \text{ °C}^{-1}$$

Contributions are at 20°C

$\rho = 579$ kg m^{-3} (at 20 °C, Appendix D)

$$k = (3.56\times10^{-5})(2.32)\left(\frac{579^4}{58}\right)^{0.33} \qquad = 0.096 \text{ W m}^{-1} \text{ °C}^{-1} \tag{8.12}$$

Gas:

Use Equation 8.13

1. Estimate the C_p using the equation and data given in Appendix D
2. Ignore the effect of pressure on C_p
3. Viscosity is given

Solution 8.5

See Section 8.9.1 and the solution to Problem 8.4.

Solution 8.6

See problem statement.

Solution 8.7

Refer to Section 8.10. Use the Watson equation. The heat of vaporisation of methyl-t-butyl ether at its boiling point is given in Appendix D.

Solution 8.8

Refer to Section 8.12.1, Equation 8.21 and Example 8.11.

Solution 8.9

Refer to Section 8.12.2, Equation 8.22 and Example 8.12.

Solution 8.10

Refer to Section 8.13, Equation 8.23 and Example 8.13.

Solution 8.11

Lydersen's method is given in Section 8.14 and illustrated in Example 8.14

Solution 8.12

Use Fig. 8.4 as an aid to selecting a suitable method.

1. hydrocarbon
2. a small amount of C_6
3. H_2 not present
4. $P > 1$ bar
5. $T < 750K$
6. $P < 200$ bar

Therefore, Use G-S

Solution 8.13

Non-ideal so use UNIQUAC equation. Check DECHEMA (1977) for binary coefficients. If not available estimate using the UNIFAC equation

Solution 9.1

For Question 1 – Toluene:

Determine the vapour pressure at 25°C. Use the Antoine equation, see Chapter 8, Section 8.11. (Coefficients from Appendix D)

$$\ln P = 16.0137 - \frac{3096.52}{298 - 53.67}$$

P = 28.22 mmHg

So, percentage toluene in the atmosphere above the liquid would be $\left(\frac{28.22}{760}\right)100 = 3.7$ %

Flammability range (Table 9.2) = 1.4 to 6.7 %.

So the concentration of toluene would be within the flammability range and a floating head or N_2 purged tank would be needed.

Solution 9.2

See Section 9.4 and the Dow guide.

It will be necessary to develop a preliminary flow-sheet for the process to determine the equipment needed and the operating conditions. An estimate will also be needed of the storage requirements.

The selection of the material factor to use is an important step.

For example, in the reaction of nitrobenzene and hydrogen, the choice will be made from the follow factors:

	Material factor
Nitrobenzene	14
Aniline	10
Hydrogen	21

Cyclo-hexylamine is only present in relatively small quantities

For this project the plant would be best split in to two sections and a Dow F&E index prepared for each. Hydrogen would be selected for the material factor for the reactor section and aniline for the separation and purification section.

Solution 9.3

HAZOP analysis is a group activity. The guide words are used to spark off discussion amongst a group of people with varied backgrounds and experience.

So, the activity has limited value when performed by students without the help of experienced engineers.

However, it is worth a group of students following through the method to gain experience of the technique.

It is important to use the guide words to generate ideas, however absurd they seem, then apply critical judgement to eliminate those that are implausible.

Solution 10.1

See section 10.4.3, Figure 10.16 and example 10.1

Solution 10.2

Section 10.4 Solid- Liquid Separation

Figure 10.10 Solid-liquid separation techniques

Solids 10%, particle size 0.1mm = $0.1 \times 10^{-3} \times 10^{6}$ = 100 microns

Possible separators: filters and centrifuges, cyclones, classifiers

Reject filters, classifiers and cyclones, material likely to be sticky and flammable

Consider centrifuge, Section 10.4, solid bowl batch or continuous likely to be most suitable for the duty.

Sigma theory

Overflow 1000 kg/h, $\rho_s = 1100$ kg/m^3, $\rho_{mix} = 860$ kg/m^3, $\mu_L = 1.7$ mNm^{-2}s , solids 10%

Density of solids is given, density of clarified liquid overflow needed. Specific volumes are additive, see chapter 8, section 7.1 .

$$1/860 = 1/\rho_L \times 0.9 + 1/1100 \times 0.1$$

$$\rho_L = 840 \text{ kg/m}^3$$

$$\Delta\rho = 1100 - 860 = 240 \text{ kg/m}^3$$

$$u_g = (240 \times (0.1 \times 10^{3})^{2}\ 9.8)/(18 \times 107\times10^{-3} = 0.000769,\ 7.7 \times 10\text{-}4 \text{ m/s} \qquad (10.2)$$

$$Q/\Sigma = 2 \times 0.000769 = 1.53 \times 10^{-3} \qquad (10.3)$$

$$Q = 1000 / 840 = 1.19 \text{ m}^3/\text{h}$$

From table 10.6 a solid bowl/basket centrifuge should be satisfactory. A Continuous discharge type should be selected. The centrifuge could be housed in a casing purged with nitrogen.

Solution 10.3

Data

Flow-rate 1200 l/m, recovery 95% greater than 100 mμ

Density of solid 2900 kgm^{-3}

Properties of water: viscosity 1300×10^{-6} Nm^{-2} s at 10 ^{o}C, 797×10^{-6} Nm^{-2} s at 30 ^{o}C;

density 999.7 kgm^{-3} at 10 ^oC, 995.6 kgm^{-3} at 30 ^oC.

Design for both temperatures.

At 10 ^oC

$$(\rho_S - \rho_L) = 2900 - 999.7 = 1900.3 \ kgm^{-3}, \ 1.9 \ gm^{-3}$$

From Fig. 10.22, for 95% recovery greater than 100 μm., d_{50} = 64 μm

From Fig. 10 23 for, a liquid viscosity of 1.3 mNm^{-2} s, D_c = 110 cm

At 30 ^oC

$$(\rho_S - \rho_L) = 2900 - 995.6 = 1904.4 \ kgm^{-3}$$

liquid viscosity = 0.8 mNm^{-2} s

D_c = 150 cm

Take the larger diameter and scale the cyclone using the proportions given in Fig. 10.24.. See section 10.4.4, example 10.2 .

Solution 10.4

See section 10.8.3 and example 10.4 .

Solution 10.5

I will treat this as a simple separation of water and acrylonitrile. In practice acrylonitrile will be soluble to some extent in water and water in acrylonitrile. Also, the azetropic composition will not be that given in the problem specification.

The design of liquid-liquid separators, decanters, is covered in section 10.6.1 and illustrated in example 10.3.

Take the acrylonitrile as the continuous phase.

Physical properties

Viscosity of acrylonitrile, estimated using the correlation given in Appendix D,

Log μ = 343.31(1/293 – 1/210.42) , μ = 0.5 mN m^{-2}s

Density of acrylonitrile = 806 kg/m^3

Water density = 998 kg/m^3

Decanter sizing

Take the droplet size as 150 μm

Then the settling velocity, $u_d = \dfrac{(150\text{x}10^{-6})^2 \text{ x } 9.8(998 - 806)}{18 \text{ x } 0.35\text{x}10^{-3}}$ (10.7)

$= 0.0067$ m/s $= 6.7 \text{ x } 10^{-3}$ m/s

Greater than $4.0 \text{ x } 10^{-3}$, so use $4.0 \text{ x } 10^{-3}$ m/s

Feed rate 300 kg/h, very small so use a vertical separator

Acrylonitrile flow-rate = 300 x 0.95 = 285 kg/h

Volumetric flow-rate, L_c = 285/(806 x 3600) = $9.82 \text{ x } 10^{-5}$ m^3/s

Continuous phase velocity must be less than the droplet settling velocity, which determines the cross-sectional area required,

$A_i = (9.82 \text{ x } 10^{-5})/(4.0 \text{ x } 10^{-3}) = 0.0246 \text{ m}^2$

So, decanter diameter = $\sqrt{[(4 \text{ x } 0.0246)/ \Pi]}$ = 0.18 m

Take height as twice diameter = 0.18 x 2 = 0.36 m

Take dispersion band as 10% of vessel height = 0.036 m

Droplet residence time = $0.036/4 \text{ x } 10^{-3}$ = 9 sec, low

The decanter is very small, due to the low flow-rate. So increase to, say, diameter 0.5 m and height 1.0 m to give a realistic size.

Check residence time for larger decanter,

Total volumetric flow = 285/(806 x 60) + 15/(998 x 60) = 0.00614 m^3/min

Volume of decanter = $1 \text{ x } 0.5^2 \text{ x } \Pi/4$ = 0.194 m3

Residence time = 0.194/0.00614 = 32 min. More than sufficient for separation, 5 to 10 minutes normally reckoned to be adequate.

Piping arrangement.

Keep velocity in feed pipe below 1 m/s.

Volumetric flow rate = 0.00614/60 = 0.000102m^3/s

area of pipe = 0.000102/1.0 = 0.000102 m^2 , diameter = $\sqrt{[(4 \times 0.000102)/\Pi]}$ = 0.0114 m = 12mm

Take the interface position as halfway up the vessel and the water take off at 90% of the height, then

$$z_1 = 0.9\text{m}, \quad z_3 = 0.5\text{m}$$

$$z_2 = (0.9 - 0.5)806/998 + 0.5 = 0.82\text{m} \qquad (10.5)$$

say 0.8m

Solution 10.6

See the solution to problem 10.5 and example 10.3

Solution 10.7

See section 10.9.2 and example 10.5

Solution 10.8

See section 10.9.2 and example 10.5

Solution 1.1

See section 11.3.2 and the dew and bubble point calculations in example 11.9.

This type of problem is best solved using a spread-sheet, see the solution to problem 11.2.

Solution 11.2

This problem has been solved using a spread-sheet (MS WORKS). The procedure set out in example 11.1 was followed.

The L/V ratio is made a variable in the spread-sheet and progressively changed until convergence between the assumed and calculated value is achieved.

i	zi	Ki	Ki.zi	zi/Ki
C3	0.05	3.3	0.165	0.02
iC4	0.15	1.8	0.27	0.08
nC4	0.25	1.3	0.325	0.19
iC5	0.2	0.7	0.14	0.29
nC5	0.35	0.5	0.175	0.70
sum	1		1.075	1.28

So feed is 2-phase.

Try L/V = 4.4

	Ki	Ai = L/VKi	Vi = Fzi/(1+Ai)
C3	3.3	1.33	2.14
iC4	1.8	2.44	4.35
nC4	1.3	3.38	5.70
iC5	0.7	6.29	2.75
nC5	0.5	8.80	3.57
	sum		18.52

L= 81.48 L/V = 4.40

convergence test % -0.02

Solution 11.3

As the relative volatility is low, this problem can be solved using the Smoker equation.

The required recovery of propylene overhead is not specified. So, the Smoker equation program was used to determine the relationship between recovery and the number of stages needed.

Mol masses propane 44. Propylene 42

$$\text{Feed composition, mol fraction propylene} = \frac{90/42}{10/44 + 90/42} = 0.904$$

The bottoms composition can be determine by a material balance on propylene:

D = overheads, B = bottoms , F = feed = 100kmol/h

$$0.904 \times 100 = 0.995D + x_b B$$

$$D + B = 100$$

Recovery, $Q = D/F = 0.995D/(100 \times 0.904)$

So, $D = 90.854Q$, $B = 100 - 90.854Q$

and $x_b = (90.4 - 90.854Q)/(100 - 90.854Q)$

Q =	0.95	0.99	0.995
x_b =	0.299	0.045	0.000028

Number of stages calculated using the Smoker equation program

Rectifying section	33	33	33
Stripping section	52	73	152
Total	85	106	185

The number of stages required increases markedly when the recovery is increased to above 0.95. The higher the recovery the lower the loss of propylene in the bottoms. The loss of revenue must be balanced against the extra cost of the column. A recovery of 0.99 would seem to be a good compromise.

Loss of propylene per 100 kmol/h feed = $0.045(100 - 90.854 \times 0.99) = 0.45$ kmol/h

Solution 11.4

Outline solution only

1. Make rough split between the tops and bottoms.

   Overheads, 98% recovery of nC4 = 270 x 0.98 = 265 kg/h

   Bottoms, 95% recovery of iC5 = 70 x 0.95 = 67 kg/h

	feed	tops	btms kg/h
C3	910	910	
i C4	180	180	
nC4	270 LK	265	5
iC5	70 HK	3	67
nC5	90		90
nC6	20		20

2. Estimate the bubble and dew points of the feed, tops and bottoms using the methods given in section 11.3.3, equations 11.5a and 11.5b. See also example 11.9.

3. Relative volatility of each component = K_i / K_{HK}. K values from the De Priester charts, section 8.16.6.

3. Determine N_m from equation 11.58.

4. Determine R_m using equations 11.60, 11.61.

5. Find N for a range of reflux ratios. Erbar and Madox method, Fig 11.1; see example 11.7.

6. Select the optimum reflux ratio.

7. Find the number of theoretical plates need at the optimum reflux.

8. Determine the feed point using the Kirkbride equation, 11.62.

9. Estimate the column efficiency using O'Connell's correlation, Fig. 11.13.
   The liquid viscosity can be estimated using the method given in Appendix D.

   The problem asks for the stage efficiency, but as a rigorous method has not been used to determine the number of theoretical plates, an estimate of the overall efficiency will be good enough. The stage (plate) efficiency could be estimated using the AIChemE method given in section 11.10.4.

10. Calculated the actual number of plates required and the feed point.

11. Estimate the column diameter using equation 11.79.

Solution 11.5

As this is a binary system, the McCabe-Thiele method described in section 11.5.2 and illustrated in example 11.2 can be used.

Compositions

Feed 60% mol acetone, overheads 99.5%mol acetone.

Material balance on 100 kmol/h feed.

Acetone overhead, 95 % recovery, = 60 x 0.95 = 57 kmol/h

Acetone in bottoms = 60 - 57 = 3 kmol/h

Total overheads = (57/99.5) x 100 = 57.3 kmol/h

Total bottoms = 100 - 57.3 = 42.7 kmol/h

Mol fraction acetone in bottoms = 3/42.7 = 0.070 (7% mol)

q line

The feed is essentially at its boiling point, 70.2 °C, so the q-line will be vertical.

McCabe-Thiele method

1. Draw the diagram using the equilibrium data given in the problem, use a large scale.
2. Determine the minimum reflux ratio.
3. Draw in the operating lines for a reflux ration 1.5 times the minimum
4. Step off the number of theoretical plates.
5. Step off the number of real plates using the plate efficiency given; see Fig.11.6.

The accuracy of the determination of the number of plates required in the rectifying section can be improved by plotting that section of the equilibrium diagram on a log scale; see example 11.2.

Solution 11.6

As this is to be treated as a binary system, the McCabe-Thiele method can be used to determine the number of theoretical stages; see section 11.5.2 and example 11.2.

The stage efficiency can be estimated using Van Winkle's correlation or the AIChemE method; see section 11.10.

The design of sieve plates is covered in section 11.13 and illustrated in example 11.11.

In practice, a side stream containing the fusel oil would be taken off a few plates from the bottom of the column. The acetaldehyde in the feed would go overhead and be recovered in a separate column.

Solution 11.7

Summary

Feed 0.9 MEK, Bottoms 0.99 Butanol, 0.01 MEK,

Feed rate 20 kmol/h, feed temperature 30 ^{o}C, boiling point 80 ^{o}C

Reflux ratio 1.5 x R_{min}.

Properties

Latent heats: MEK 31284 kJ/kmol, 2-butanol 40821 kJ/kmol

Specific heats: MEK 164 kJ/kmol, 2-butanol 228 kJ/kmol

Mol mass: MEK 72.11, 2-butanol 74.12

Solution

(a), (b) minimum reflux ratio and number of theoretical stages

Binary system, so use McCabe-Thiele method to find the minimum reflux ratio and number of stages; see example 11.2.

Latent heat of feed = 0.9 x 31234 + 0.1 x 40821 = 32,193 kJ/kmol

Sensible heat to bring feed to boiling point = (0.9 x 164 + 0.1 x 228)(80 – 35)

= 7668 kJ/kmol

q = (32193 + 7668) / 32193 = 1.24

Slope of q line = 1.24 / (1.24 – 1) = 5.2

From McCabe-Thiele plot ϕ_{min} = 0.66

R_{min} = (0.99/0.66) - 1 = 0.5 (11.24)

R = 1.5 x 0.5 = 0.75, ϕ = 0.99 / (1 + 0.75) = 0.57

For this reflux ratio, stepping off the stages on the McCabe-Thiele diagram gives

8 stages below the feed and 8 above, total 16 theoretical stages.

The diagram was enlarged by a factor of 8 above the feed to accurately determine the number of stages.

(*c*) Plate efficiency

The question asked for the stage efficiency. I will estimate the overall column efficiency using O'Connell's correlation. The individual stage efficiency could be estimated, after the designing the plates, using Van Winkle's correlation, (11.69) or the AIChemE method, section 11.10.4.

Liquid viscosity's at the average column temperature:

MEK 0.038 Nm^{-2} s, Butanol 0.075 Nm^{-2} s

μ_a at feed composition = 0.9 x 0.038 + 0.1 x 0.075 = 0.042 Nm^{-2} s

α_a = 2.6 (α can be estimated from the equilibrium data using (11.23)).

μ_a x α_a = 0.042 x 2.6 = 0.109

E_o = 51 – 32.5 x Log (0.109) = 82.3 % (11.67)

Seems rather high, would need to confirm before use in practice.

Table 11.2 gives a value for Toluene – MEK as 85 %. So use 80 % for the remainder of the question.

(d) number of actual stages

Number of real plates = 16/0.8 = 20

(e) plate design

Flow-rates

Feed = 20kmol/h

Mass balance on MEK, $0.9 \times 20 = 0.99 D + 0.1B$

Overall balance, $20 = D + B$,

which gives $D = 18.16$ and $B = 1.40$ kmol/h

From the McCabe-Thiele diagram the slope of the bottom operation line, (L_m' / V_m')

$$= 0.95/0.90 = 1.056$$

$V_m' = L_m' - B$, so $L_m' / (L_m' - 1.40) = 1.056$ hence,

$L_m' = 1.056 L_m' - 1.40 \times 1.056, \quad = 1.478 / 0.056 = 26.12$ kmol/h

$V_m' = 26.12 - 1.40 = 24.72$ kmol/h

Densities

2-butanol, at feed temperature, 80 °C, = 748 kgm^{-3},
at bottoms temperature, 99.5 °C, = 725 kgm^{-3}.

The properties of MEK will be very similar, so ignore the change in composition up the column.

Design for conditions at the base.

Base pressure

Allow 100 mm WG per plate. Number of plates, allowing for reboiler = 19

$\Delta P = 19 \times 100 = 1900$ mm WG $= 1.9 \times 1000 \times 9.8 = 18620$ N/m^2

Say, allow 19 kN/m^2, column base pressure = 100 + 19 = 119 kN/m^2

$$\text{Vapour density} = \frac{74.12}{22.4} \times \frac{273}{(273 + 99.5)} \times \frac{119}{100} = 2.9 \text{ kgm}^{-3}$$

Column diameter

$L_m'/ V_m' \cong L_w'/ V_w' = 1.056$

$$F_{Lv} = 1.056 \sqrt{(2.9 / 725)} = 0.067 \qquad (11.82)$$

The column diameter is likely to be small, as the feed rate is low, so take the plate spacing as 0.45 m.

From Fig. 11.27, K_1 = 0.078

Surface tension, estimated using (8.23), = 9.6 mJ/m^2 (mN m).

Correction = $(0.0096/0.02)^{0.2}$ = 0.86

Corrected K_1 = 0.078 x 0.86 = 0.067

u_f = $0.067 \sqrt{(725 - 2.90)/2.90)}$ = 1.06 m/s

Take design velocity as 80% of flooding,

Maximum velocity = 1.06 x 0.8 = 0.85 m/s.

Volumetric flow-rate = (24.72 x 74.12)/(2.90 x 3600) = 0.176 m^3/s

So, area required = 0.176 / 0.85 = 0.21 m^2

Take downcomer area as 12 %, then minimum column area required

= 0.21 / (1 – 0.12) = 0.24 m^2

Column Diameter = $\sqrt{(4 \times 0.24) \Pi}$ = 0.55 m

Liquid flow pattern

Max. vol. liquid flow-rate = 26.12 x 74.12)/(725 x 3600) = 0.74×10^{-3} m^3 /s

Fig. 11.28, column diameter is off the scale but liquid rate is low so try a reverse flow plate. Adapt design method for across-flow plate

Keep downcomer area as 12% , A_d / A_c = 0.12

From Fig. 11.31, l_w / D_c = 0.76

Take this chord for the reverse flow design.

Then down comer width, weir length = (0.76x 0.55) / 2 = 0.21 m

Summary, provisional plate design

Column diameter = 0.55 m

Column area, A_c = 0.23 m^2

Downcomer area, A_d = 0.06 x 0.23 = 0.014 m^2

Net area, A_n = A_c - A_d = 0.23 - 0.014 = 0.216 m^2

Active area, A_a = A_c - 2 A_d = 0.202 m^2

Hole area, take as 10% of A_a , A_h = 0.02 m^2

As column diameter is small and liquid flow-rate low, take weir height as 40 mm, plate thickness 4 mm. and hole dia. 5 mm.

Check on weeping

Design at rates given in the question, for illustration; turn down ratio not specified.

Liquid rate = (26.12 x 74.12) / 3600 = 0.54 kg/s

h_{ow} = $750(0.54 / (725 \times 0.21))^{2/3}$ 17.5 mm (11.85)

. h_{ow} + . h_w = 17.5 + 40 = 57.5

From Fig 1130, K_2 = 33

$u_{h\,min}$ = $[33 - 0.90\,(25.4 – 5)] / (2.9)^{1/2}$ = 8.5 m/s (11.84)

Vapour rate = 0.176 m^3/s, so velocity through holes, u_h =

0.176 / 0.02 = 8.8 m/s

Just above weep rate. Need to reduce hole area to allow for lower rates in operation.

Try 8% , A_h = 0.0202 x 0.08 = 0.014

u_h = 0.176 / 0.014 = 12.6 m/s - satisfactory

Plate pressure drop

Plate thickness / hole diameter = 4/5 = 0.8

From Fig. 11.34, C_o = 0.77

h_d = $51(12.6 / 0.77)^2 \times (2.9 / 725)$ = 54.6 mm (11.88)

h_r = 12.5 x 103 / 725 = 17.3 mm (11.89)

h_t = 54.6 + 57.5 + 17.3 = 129.4 mm (11.90)

Downcomer liquid back-up

Head loss under downcomer

Take h_{ap} at 5 mm below the top of the weir (see Fig. 11.35)

Then $A_{ap} = 0.21 \times (40 - 5) = 0.00735 \text{ m}^2$

So, $A_{ap} < A_d$ and $A_m = A_{ap} = 7.35 \times 10^{-3} \text{ m}^2$

$h_{dc} = 166 [0.54 / (725 \times 7.35 \times 10^{-3})]^2 = 1.7$ mm (11.92)

Back-up, $h_b = 57.5 + 129.4 + 1.7 = 188.6$ mm

Which is less than half the plate spacing plus the weir height, so the design is satisfactory.

Check residence time

$t_r = (0.014 \times 188.6 \times 10^{-3} \times 725) / 0.54 = 3.5$s, acceptable

Check entrainment

u_v = Vol. Flow-rate / net column area = 0.176 / 0.216 = 0.815 m/s

u_f (flooding vel.) = 1.06 m/s, so percent flooding = 0.815 / 1.06 = 77%

F_{Lv} = 0.067 (calculated previously)

From Fig. 11.29, $\psi = 3.5 \times 10^{-2}$, satisfactory.

Conclusion

Design using cross flow plates looks feasible.

Solution 11.8

The number of theoretical stages can be determined using the McCabe-Thiele method illustrated in example 11.2, section 11.5.

For the plate column, the column efficiency can be approximated using the value given in Table 11.2. The column diameter can be estimated using equation 11.79.

For the packed column, the HETP value given in Table 11.4 can be used to estimate the column height. The column diameter can be calculated using the procedure given in section 11.14.4.

Having sized the columns, the capital costs can be compared using the procedures and cost data given in Chapter 6.

The column auxiliaries and operation costs are likely to be more or less the same for both designs

Solution 11.9

See section 11.16.2, example 11.15.

Feed 2000 kg/h, 30 % MEK

Solvent 700 kg/hr, pure TCE

MEK in feed = 600 x 0.3 = 600 kg/h

Water in raffinate = 2000 – 600 = 1400 kg/hr

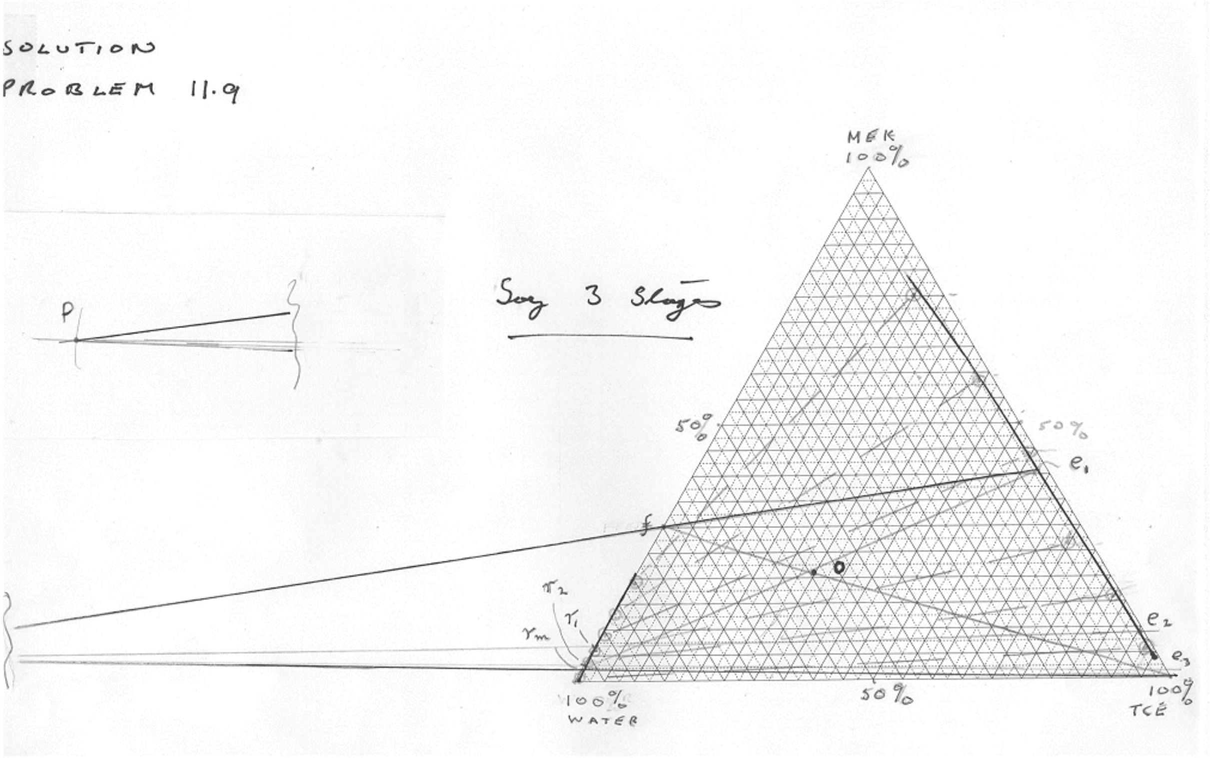
At 95% recovery, MEK in final raffinate = (1 – 0.95) x 600 = 30 kg/hr

Composition at the point o = (600) / (2000 + 700) = 0.22 MEK, 22%

Composition of final raffinate = 30 / (1400 + 30) = 0.21 MEK, 2.1%

Following the construction set out in section 11.16.2 gives 3 stages required, see diagram.

Diagram



Solution 11.10

See section 11.14 and example 11.14.

For this design, as the solution exerts no back-pressure the number of overall gas phase transfer units can be calculated directly from equations 11.107 and 11.108.

$$\Delta y = y$$

so, $y_{lm} = (y_1 - y_2)/ \ln(y_1/y_2)$ (11.108.)

and, $N_{OG} = (y_1 - y_2)/ y_{lm} = \ln(y_1/y_2)$ (11.107)

When estimating the height of an overall gas phase transfer unit, note that as there is no back pressure from the liquid the slope of the equilibrium line, *m*, will be zero; i.e. there is no resistance to mass transfer in the liquid phase.

Ceramic or plastics packing would be the most suitable this column.

Solution 13.1

See section 13.3.4, equations 13.7 to 13.18

Solution 13.2

Use equation 13.34 .

(a) rigid constant C = 0.43

(b) free to rotate, C = 0.56

Solution 13.3

See section 13.5.1

Use equation 13.39 for the cylindrical section and equation 13.40 or 13.41 for the ends.

Solution 13.4

Specification

Shell 387 mm id, tubes 14.83 mm id, 19.05 mm od, length 6096 mm

Kerosene in the shell, operating pressure 5 bar.

Crude in the tubes, operating pressure 6.5 bar

Material of construction, semi-killed or silicon killed carbon steel.

(a) Design pressures: take as 10% greater than operating pressures; section 13.4.1.

Shell = (5 – 1) x 1.1 = 4.4 bar = 4.4 x 10^5 N/m^2
Tubes = (6.5 – 1) x 1.1 = 6.05 bar = 6.5 x 10^5 N/$m^{2.}$

Design temperature: maximum operating temperature = 200 °C. Take this as the design temperature for both the shell and tubes. The tubes could reach the kerosene temperature if there was no flow of crude oil; section 13.4.2.

(b) Corrosion allowance: no information is given on the purity of the kerosene or the composition of the crude. If sulphur free the kerosene should not corrode. If wet the crude could be corrosive.

Take the kerosene allowance as 2 mm and the crude as 4 mm; section 13.4.6

(c) End covers: shell and floating head use torispherical, header- cover flat plate.; see figure, example 12.2.

(d) Stressing: take design stress as 105 N/mm^2 at 200 °C; , Table 13.2.

Shell: $e = \dfrac{4.4 \times 10^5 \times 0.387}{2 \times 105 \times 10^6 - 4.4 \times 10^5} = 0.0008 \text{ m} = 0.8 \text{ mm}$ (13.39)

add corrosion allowance = 0.8 + 2 = 2.8 mm

This is less than the minimum recommended thickness, section 13.4.8, so round up to 5 mm.

Header: $e = \dfrac{6.05 \times 10^5 \times 0.387}{2 \times 105 \times 10^6 - 6.05 \times 10^5} = 0.0011 \text{ m} = 1.1 \text{ mm}$ (13.39)

add corrosion allowance = 1.1 + 4 = 5.1 mm

Shell end-cover, torispherical, take $R_c = 0.3$, $R_k/R_c = 0.1$; section 13.5.4

$C_s = \frac{1}{4}(3 + \sqrt{10}) = 2.37$. Take joint factor as = 1.0, formed head.

$e = \dfrac{4.4 \times 10^5 \times 0.3 \times 2.37}{2 \times 105 \times 10^6 \times 1 + 4.4 \times 10^5 (2.37 - 0.2)} = 0.00148 \text{ m} = 1.5 \text{ mm}$ (13.44)

add corrosion allowance = 1.5 = 2 = 3.5 mm

Floating-head cover, torispherical:

Bundle to shell clearance, Fig 12.10 ≅ 53 mm, take as split ring.

$D_b = 0.387 - 0.334 = 0.334$ mm

Take R_c as 0.3, $R_k/R_c = 0.1$

$C_s = 2.37$

$e = \dfrac{6.05 \times 10^5 \times 0.3 \times 2.37}{2 \times 105 \times 10^6 \times 1 + 6.05 \times 10^5 (2.37 - 0.2)} = 0.00206 \text{ m} = 2.1 \text{ mm}$ (13.44)

add corrosion allowance = 2.1 + 4 = 6.1 mm

Flat plate (header cover):

Type (e) Fig 13.9, Cp = 0.55. D_i = 387 mm, so $D_e \cong 0.4$ m

$e = 0.55 \times 0.4 \sqrt{(6.05 \times 10^5 / 105 \times 10^6)} = 0.167 \text{ m}$ (13.42) add corrosion allowance = 16.7 + 4 = 20.7 mm

All thicknesses would be rounded to nearest standard size.

(e) Tube rating

Tube id = 14.83 mm, od = 19.05 mm, design stress 105 x 10^6 N/m^2,
design pressure 6.05 N/m^2.

Thickness required, $e = \dfrac{6.05 \times 10^5 \times 1.83 \times 10^{-3}}{2 \times 105 \times 10^6 - 6.05 \times 10^5} = 0.0000053$ m (13.39)

= 0.005 mm

Actual wall thickness = (19.05 - 14.83)/2 = 2.1 mm.

So ample margin for corrosion.

(f) Tube-sheet thickness should not be less than tube diameter; section 12.5.8.

So take thickness as = 20 mm

(g) Would use weld neck flanges; Appendix F.

Shell od = 387 + (2 x 5) = 397 mm, say, 400 mm

Design pressure = 4.05 x 10^5 N/m^2, design temperature = 200 °C

6 bar rating would be satisfactory, table 13.5.

Floating head od ≅ 350 mm, design pressure = 6.05 x 10^5 N/m^2, design temperature 200 °C.

Use a 10 bar rated flange, table 13.5.

(h) Supports

Use saddle supports, section 13.9.1, Fig 13.26*d*.

Smallest size given in Fig 13.26*d* is 600 mm diameter. So, scale all dimensions to 400 mm and make all plate 5 mm.

Rough check on weight

Diameter ≅ 0.4 m, length ≅ 10 m

Shell and header, volume of steel = $\Pi \times 0.4 \times 10 \times x \times 10^{-3} = 0.063\ m^3$

Volume of shell head, take as flat, $\cong$ $(\Pi/4 \times 0.4^2 \times 3.5 \times 10^{-3}) = 0.0004\ m^3$

Volume of floating head, take as flat $\cong$ $(\Pi/4 \times 0.334^2 \times 6 \times 10^{-3}) = 0.0005\ m^3$

Volume of flat plate end cover $= \Pi/4 \times 0.4^2 \times 21 \times 10^{-3} = 0.0026\ m^3$

Volume of tube-sheet $= 0.0026\ m^3$, ignoring the holes

Volume of tubes $= 168 \times [\Pi/4\ (19.05^2 - 14.83^2) \times 10^{-6} \times 6.09] = 0.115\ m^3$

Number of baffles $= 6090/77.9 - 1 = 77$

Taking baffles as 3 mm thick and ignore the baffle cut,

volume $= 77(\Pi/4 \times 0.387^2 \times 3 \times 10^{-3}) = 0.027\ m^3$

Total volume of steel

Shell	0.063
Shell head	0.0004
Floating head	0.0005
End-cover	0.0026
Tube-sheet	0.0026
Tubes	0.115
Baffles	0.027
Total	$0.21\ m^3$

Taking density of steel as $7800\ kg/m^3$, mass of exchanger $= 0.21 \times 7800 = 1638$ kg

Weight $= 1638 \times 9.8 = 16{,}052$ N $= 16$ kN

Mass of water, ignore volume of tubes, $= 1000(\Pi/4 \times 0.4^2 \times 10) = 1257\ m^3$

Weight $= 1257 \times 9.8 = 12319$ N $= 12$ kN

Maximum load on supports $= 16 + 12 = 28$ kN

Load given in Table 13.26*a* for a 600 mm diameter vessel = 35 kN, so design should be satisfactory.

Solution 13.5

The design procedure will follow that set out in solution 13.4.

The exchanger will be hung from brackets, see section 13.9.3.

Solution 13.6

The procedure for solving this problem follows that used in examples 13.3 and 13.4.

1. Determine the minimum plate thickness to resist the internal pressure, equation 13.39.

2. Select and size the vessel ends, use torispherical or ellipsoidal heads; section 13.5.4

3. Increase the basic plate thickness to allow for the bending stress induced by the wind loading at the base of the vessel, and the small increase in stress due to the dead weight of the vessel.

4. Check that the maximum combined stresses at the base are within the design stress and that the critical buckling stress is not exceeded.

5. Decide which openings need compensation. The 50 mm nozzles are unlikely to need compensation but the vapour outlet and access ports probably will. Use the equal area method for determining the compensation required; section 13.6.

6. Use standard flanges; section 13.10.5 and appendix F.

7. Design the skirt support. A straight skirt should be satisfactory. Consider the wind load, the weight of the vessel, and the weight of the vessel full of water. Though the vessel is not likely to be pressure tested during a storm, a fault condition could occur during operation and the vessel fill with process fluid. The process fluid is unlikely to be more dense than water.

8. Design the base ring following the method given in section 13.9.2.

Solution 13.7

Only the jacketed section need be considered, the vessel operates at atmospheric pressure.

The jacketed section of the vessel will be subjected to an external pressure equal to the steam pressure (gauge).

The jacket will be under the internal pressure of the steam.

Operating pressure $= 20 - 1 = 19$ barg $= 19 \times 10^5$ N/m^2 $= 1.9$ N/mm^2

o.d. of vessel $= 2 + 2 \times 25 \times 10^{-3} = 2.05$ m

i.d. of jacket $= 2.05 + 2 \times 75 \times 10^{-3} = 2.2$ m

Jacket, required thickness, $e = \dfrac{1.9 \times 2.2}{2 \times 100 - 1.9} = 0.021$ m $= 21$ mm (13.39)

So the specified thickness of 25 mm should be OK, with adequate margin of safety.

Vessel section:

Take Poisson's ratio, ν, for carbon steel as 0.3.
E is given as 180,000 $N/mm^2 = 1.8 \times 10^{11} N/m^2$

Check collapse pressure without any consideration of stiffening

$$P_c = 2.2 \times 1.8 \times 10^{11} (25 \times 10^{-3}/2.05)^3 = 718{,}214 \text{ N/m}^2 = 7.2 \text{ bar} \qquad (13.51)$$

So vessel thickness is adequate to resist the steam pressure.

Solution 13.8

The pipe is a thick cylinder, see section 13.15.1 and the solution to problem 13.7.

Solution 13.9

Tank diameter = 6 m, height of liquid, H_L = 16 m , density of liquid. ρ_L = 1520 kg/m^3,
g = 9.81, design stress, f_t, = 90 N/mm. Take joint factor, J, as 0.7, a safe value.

$$e_s = \frac{1520 \times 16 \times 9.81 \times 6}{2 \times 90 \times 10^6 \times 0.7} = 0.0114 \text{ m} \qquad (13.130)$$

Say 12 mm

Solution 12.1

The procedure will follow that used in the solution to problem 12.2.

As the cooling water flow-rate will be around half that of the caustic solution, it will be best to put the cooling water through the tubes and the solution through the annular jacket.

The jacket heat transfer coefficient can be estimated by using the hydraulic mean diameter in equation 12.11.

Solution 12.2

Heat balance

$$Q \ + m\ C_p\ (T_{out} - T_{in})$$

$$Q \ = \ (6000/3600) \times 4.93 \times (65 - 15) \ = \ 411 \text{ kW}$$

Cross-section of pipe = $(\Pi/4)(50 \times 10^{-3})^2 \ = \ 1.963 \times 10^{-3}$ m

Fluid velocity, $u \ = \ \frac{6000}{3600} \times \frac{1}{866} \times \frac{1}{1.963 \times 10^{-3}} \ = \ 0.98$ m/s

$$Re \ = \ \frac{866 \times 0.98 \times (50 \times 10^{-3})}{0.44 \times 10^{-3}} \ = \ 96{,}441$$

$$Pr \ = \ \frac{4.3 \times 10^{-3} \times 0.44 \times 10^{-3}}{0.3895} \ = \ 4.86$$

Liquid is not viscous and flow is turbulent, so use eqn 12.11, with C = 0.023 and neglect the viscosity correction factor.

$$Nu \ = \ 0.023(96441)^{0.8}(4.86)^{0.33} \ = \ 376$$

$$h \ = \ (0.385/50 \times 10^{-3}) \times 376 \ = \ 2895 \text{ Wm}^{-2}\,^{\circ}\text{C}^{-1}$$

Take the steam coefficient as 8000 $\text{Wm}^{-2}\,^{\circ}\text{C}^{-1}$

$$1/U_o \ = \ \frac{1}{8000} + \frac{60 \times 10^{-3}(60/50)}{2 \times 480} + \frac{60}{50} \times \frac{1}{2895} \qquad (12.2)$$

$$U_o \ = \ 1627 \text{ Wm}^{-2}\,^{\circ}\text{C}^{-1}$$

$$\Delta T_{lm} \ = \ (85 - 35)/\text{Ln}\,(85/35) \ = \ 56.4\ ^{\circ}\text{C} \qquad (12.14)$$

$$A_o \ = \ (411 \times 10^3)/(1627 \times 56.4) \ = \ 4.5 \text{ m}^2 \qquad (12.1)$$

$A_o = \Pi \times d_o \times L$, $L = 4.5/(\Pi \times 60 \times 10^{-3}) = 23.87$ m

Number of lengths = 23.87/ 3 = 8 (rounded up)

Check on viscosity correction

Heat flux, $q = 411/4.5 = 91.3$ kW/m^2

ΔT across boundary layer = $q/h = 91{,}300/2895 = 32$ °C

Mean wall temperature = $(15 + 65)/2 + 32 = 72$ °C

From table, $\mu_w \cong 300$ mN m^{-2} s

$\mu/\mu_w = (0.44/0.3)^{0.14} = 1.055$, so correction would increase the coefficient and reduce the area required.

Leave estimate at 8 lengths to allow for fouling.

Solution 12.3

Physical properties. from tables

Steam temperature at 2.7 bar = 130 °C

Mean water temperature = $(10 + 70)/2 = 40$ °C

Density = 992.2 kg/m^3, specific heat = 4.179 kJ kg^{-1} °C^{-1}, viscosity = 651×10^{-3} N m^{-2} s, Thermal conductivity = 0.632×10^{-3} W m^{-1} °C^{-1}, Pr = 4.30.

Take the material of construction as carbon steel, which would be suitable for uncontaminated water and steam, thermal conductivity 50 W m^{-1} °C^{-1}.

Try water on the tube side.

Cross–sectional area = $124 (\Pi/4 \times (15 \times 10^{-3})^2) = 0.0219$ m^2

$$\text{Velocity} = \frac{50000}{3600} \times \frac{1}{992.2} \times \frac{1}{0.0219} = 0.64 \text{ m/s}$$

$$Re = \frac{992.2 \times 0.64 \times 15 \times 10^{-3}}{0.651 \times 10^{-3}} = 14{,}632 \qquad (1.5 \times 10^{-4})$$

From Fig 12.23, $j_h = 4 \times 10^{-3}$

$Nu = 4 \times 10^{-3} \times 14632 \times 4.0^{-0.33} = 92.5$

$h_i = 92.5 \times (632 \times 10^{-3}) / 15 \times 10^{-3} = 3897\ Wm^{-2}\,°C^{-1}$

Allow a fouling factor of 0.0003 on the waterside and take the condensing steam coefficient as 8000 $Wm^{-2}\,°C^{-1}$; see section 12.4 and 12.10.5.

$$1/U_o = (1/3897 + 0.0003)(19/15) + \frac{19\times10^{-3} Ln(19/15)}{2 \times 50} + 1/8000 = .000875$$

$U_o = 1143\ Wm^{-2}\,°C^{-1}$

$$\Delta T_{lm} = \frac{(130 - 70) - (130 - 10)}{Ln\ (60/120)} =_{-} 86.6\ °C \qquad (12.4)$$

The temperature correction factor, F_t , is not needed as the steam is at a constant temperature.

Duty, $Q = (50{,}000/3600) \times 4.179(70 - 10) = 3482.5$ kW

$$\text{Area required, } A_o = \frac{3482.5 \times 10^3}{1143 \times 86.6} = 35.2\ m^2$$

Area available $= 124(\Pi \times 19 \times 10^{-3} \times 4094 \times 10^{-3}) = 30.3\ m^2$

So the exchanger would not meet the duty, with the water in the tubes.

Try putting the water in the shell.

$$\text{Flow area, } A_s = \frac{(24 - 19)}{24} 337 \times 10^{-3} \times 106 \times 10^{-3} = 7.44 \times 10^{-3}\ m^2 \qquad (12.21)$$

Hydraulic mean diameter, $d_e = (1.10/19)(24^2 - 0.917 \times 19^2) = 14.2$ mm (12.2)

Velocity, $u_s = 50000/3600 \times 1/992.2 \times 1/7.44 \times 10^{-3} = 1.88$ m/s

$$Re = \frac{992.2 \times 1.88 \times 14.2 \times 10^{-3}}{0.65 \times 10^{-3}} = 40{,}750 \quad (4.1 \times 10^4)$$

From Fig 12.29 for 25% baffle cut, $j_h = 3.0 \times 10^{-3}$

$Nu = 3.0 \times 10^{-3} \times 40750 \times 4.3^{0.33} = 198$

$h_s = 198 \times 632 \times 10^{-3}/14.2 \times 10^{-3} = 8812\ Wm^{-2}\,°C^{-1}$

A considerable improvement on the coefficient with the water in the tubes.

$$1/U_o = (1/8000)(19/15) + \frac{19\times10^{-3} Ln(19/15)}{2 \times 50} + (1/8812 + 0.0003)$$

$U_o = 1621 \text{ Wm}^{-2} \text{°C}^{-1}$

$$A_o = \frac{3482.5 \times 10^3}{1621 \times 86.6} = 24.80 \text{ m}^2$$

So the exchanger should be capable of fulfilling the duty required, providing the water in put through the shell.

Note; the viscosity correction factor has been neglected when estimating the heat transfer coefficients. Water is not a viscous liquid, sot he correction would be small.

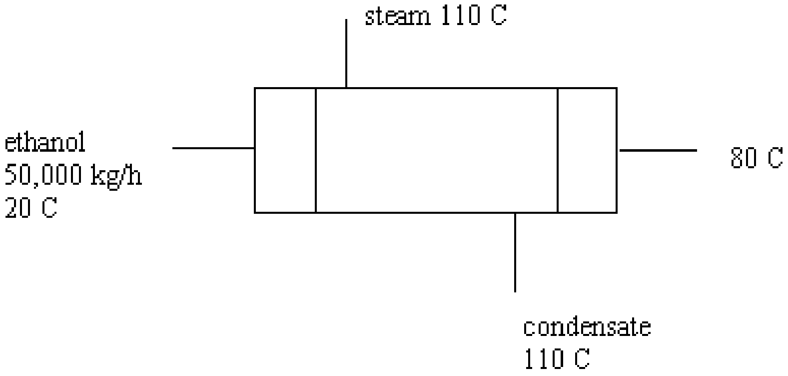
In practice, it would be necessary to check that the pressure drop on the water-side could be met by the supply pressure

Solution 12.4

There is no unique solution to a design problem. The possible solutions for this design have been constrained by specifying the tube dimensions and the disposition of the fluid streams. Specifying steam as the heating medium and putting in the shell simplifies the calculations. It avoids the need to make tedious, and uncertain, calculations to estimate the shell-side coefficient.

The heat exchanger design procedure set out in Fig. A, page 680, will be followed.

Step 1 Specification



Flow-rate of ethanol = 50000/3600 = 13.89 kg/s

Ethanol mean temperature = (20 + 80)/2 = 50 °C

Mean specific heat = $2.68 \text{ kJ kg}^{-1} \text{°C}^{-1}$ (see table step 2)

Duty = $m C_p (T_1 - T_2)$ = 13.89 x 2.68 x (80 – 20) = 2236 kW

Step 2 Physical properties

Saturation temperature steam at 1.5 bar, from steam tables, = 111.4 °C

Thermal conductivity of carbon steel = 50 W m^{-1} $°C^{-1}$

Properties of ethanol

Temp °C	$C_{p,}$ kJ kg^{-1} $°C^{-1}$	k, W m^{-1} $°C^{-1}$	ρ, kg/m^3	μ, N m^{-2} s x 10^3
20	2.39	0.164	789.0	1.200
30	2.48	0.162	780.7	0.983
40	2.58	0.160	772.1	0.815
50 (mean)	2.68	0.158	763.2	0.684
60	2.80	0.155	754.1	0.578
70	2.92	0.153	744.6	0.495
80	3.04	0.151	734.7	0.427
90	3.17	0.149	724.5	0.371
100	3.31	0.147	719.7	0.324
110	3.44	0.145	702.4	0.284

Step 3 Overall coefficient

Ethanol is not a viscous fluid, viscosity similar to water, so take a initial value for U of 1000 Wm^{-2} $°C^{-1}$, based on the values given in Table 12.1 and Fig. 12.1.

Step 4 Passes and LMTD

A typical value for the tube velocity will be 1 to 2 m/s; see section 12.7.2.

Use 1 m/s to avoid the possibility of exceeding the pressure drop specification.

Fixing the tube-side velocity will fix the number of passes; see step 7.

$$\Delta T_{lm} = \frac{(111.4 - 80) - (111.4 - 20)}{\text{Ln}((111.4 - 20)/(111.4 - 20))} = 56.16\ °C \qquad (12.4)$$

Step 5 Area

Trial area, A = $(2236 \times 10^3)/(1000 \times 56.16) = 39.8\ m^2$ (12..1)

Step 6 Type

As the mean temperature difference between the shell and tubes is less than 80 °C, a fixed tube sheet exchanger can be used.

Ethanol in the tubes, as specified.

Step 7 Number of tubes

Surface area of one tube = $\Pi \times (29 \times 10^{-3}) \times 4 = 0.364\ m^2$ (based on the o.d.)

Number of tubes needed = 39.8/0.364 = 109.3, say 110

Cross-sectional area of one tube = $\Pi/4 \times (25 \times 10^{-3})^2 = 4.91 \times 10^{-4}\ m^2$

Volumetric flow-rate of ethanol = 13.89/763.2 = $0.0182\ m^3/s$

Tube-side velocity = volumetric flow/cross-sectional area per pass

So, cross-sectional area per pass = 0.0182/1 = $0.0182\ m^2$

Number of passes = total cross-sectional area/ cross-sectional area per pass

$$= (110 \times 4.91 \times 10^{-4})/0.0182 = 2.9$$

Take as 4 passes. This will increase the tube-side velocity to above the chosen value. So, increase the number of tubes to 120, giving a uniform 30 tubes per pass. Use E type shell.

Step 8 Shell diameter

The shell diameter is not needed at this point as the shell-side coefficient is not dependent on the diameter. Leave till after checking the overall coefficient and tube-side pressure drop.

Step 9 Tube-side coefficient

Velocity, u_t = volumetric flow-rate/cross-sectional area per pass

$$= (0.0182)/(30 \times 4.91 \times 10^{-4}) = 1.24\ m/s$$

$$Re = \frac{763.2 \times 1.24 \times 25 \times 10^{-3}}{0.684} = 34,589 \quad (3.6 \times 10^4)$$

From Fig. 12.23, $j_h = 3.4 \times 10^{-3}$

$$Pr = \frac{2.68 \times 10^3 \times 0.684 \times 10^{-3}}{0.158} = 11.6$$

$$Nu = 3.4 \times 10^{-3}(34589)(11.6)^{0.33} = 264 \qquad (12.5)$$

$$h_i = (264 \times 0.158)/(25 \times 10^{-3}) = 1668\ Wm^{-2}\,°C^{-1}$$

The viscosity correction factor has been neglected as ethanol is not viscous.

Step 10 Shell-side coefficient

Take the shell-side coefficient for condensing steam as 8000 $Wm^{-2}\,°C^{-1}$; section 12.10.5
This includes the fouling factor.

Step 11 Overall coefficient

Ethanol should not foul the tubes, so take the fouling factor for the tube-side as 0.0002, that for light organics in Table 12.2.

$$1/U_o = (1/1668 + 0.0002)(29/25) + \frac{29 \times 10^{-3}(29/25)}{2 \times 50} + 1/8000 = 0.001389$$

$$U_o = 720 \text{ Wm}^{-2}\,°\text{C}^{-1}$$

Too low, so back to Step 3. Put the overall coefficient = 720 $\text{Wm}^{-2}\,°\text{C}^{-1}$

Area required = $(2236 \times 10^3)/(720 \times 56.16) = 52.3 \text{ m}^2$

Number of tubes = 52.3/0.364 =   143.7 (144)

Try 144 tubes with 4 passes.

New tube velocity = 1.24 x 120/144 =   1.03 m/s

New Re =   34589 x 1.03/1.24 = 28,731 (2.9×10^4)

From Fig 12.23, $j_h = 3.8 \times 10^{-3}$

$$Nu = 3.8 \times 10^{-3}(28731)(11.6)^{0.33} = 245$$

$$h_i = (245 \times 0.158)/(25 \times 10^{-3}) = 1548 \text{ Wm}^{-2}\,°\text{C}^{-1}$$

$$1/U_o = (1/1548 + 0.0002)(29/25) + \frac{29 \times 10^{-3}(29/25)}{2 \times 50} + 1/8000 = 0.001443$$

$$U_o = 693 \text{ Wm}^{-2}\,°\text{C}^{-1}$$

Still too low but check pressure drop with this arrangement to see if the number of passes could be increased, rather than the number tubes.

Step 12 Pressure drops

$$\Delta P_i = 4(8 \times 3.7 \times 10^{-3} \times (4/25 \times 10^{-3}) + 2..5)(763.2 \times 1.03^2/2) = 11,718 \text{ N/m}^2 \qquad (12..30)$$

$$= 0.12 \text{ bar}$$

Well below the allowable drop of 0.7 bar. So, try six passes, 24 tubes per pass.

New tube-side velocity = 1.03 x 6/4 = 1.54 m/s

New Re   = 28731 x 1.54/1.03 = 42,957 (4.3×10^4)

From Fig. 12.24 $j_f = 3.3 \times 10^{-3}$

$$\Delta P_i = 4(8 \times 3.3 \times 10^{-3} \times (4/25 \times 10^{-3}) + 2..5)(763.2 \times 1.54^2/2) = 24,341 \text{ N/m}^2$$

$= 0.24$ bar

Check on nozzle pressure drops.

Take nozzle / pipe velocity to be 2 m/s; see chapter 5, section 5.6.

Area of nozzle = volumetric flow-rate/velocity = $0.0182/2 = 0.0091 \text{ m}^2$

Nozzle diameter = $\sqrt{(4 \times 0.0091/\Pi)} = 0.108$ m

Select standard pipe size, 100 mm

Nozzle velocity = $2 \times (108/100)^2 = 2.33$ m/s

Velocity head = $u^2/2g = 2.33^2 / 2 \times 9.8 = 0.277$ m

Allow one velocity head for inlet nozzle and a half for the outlet; see section 12.8.2.

Pressure drop over nozzles = $\rho gh = 763.2 \times 9.8 \times (1.5 \times 0.277) = 3{,}108 \text{ N/m}^2$

$= 0.03$ bar

Total tube-side pressure drop = 0.24 + 0.0 3 = 0.27 bar, well below the 0.7 bar allowed .

No limiting pressure drop is specified for the shell-side.

Back to steps 9 to 11

From Fig 12.3, j_h , at Re = 4.3×10^4, = 3.2×10^{-3}

$Nu = 3.2 \times 10^{-3}(42957)(11.6)^{0.33} = 309$

$h_i = (309 \times 0.158)/(25 \times 10^{-3}) = 1{,}953 \text{ Wm}^{-2}\,°\text{C}^{-1}$

$$1/U_o = (1/1953 + 0.0002)(29/25) + \frac{29 \times 10^{-3}(29/25)}{2 \times 50} + 1/8000 = 0.001287$$

$$U_o = 777 \text{ Wm}^{-2}\,°\text{C}^{-1}$$

Greater than the assumed value of 720 $\text{Wm}^{-2}\,°\text{C}^{-1}$, so the design is satisfactory.

Shell-side design (Step 10)

Use a square pitch as high shell-side velocities are not rquired with a condensing vapour.

Take the tube pitch = 1.25 x tube o.d. = $29 \times 10^{-3} \times 1.25 = 36.25 \times 10^{-3}$ m

Bundle diameter, from Table 12.4, for 6 passes , square pitch , $K_1 = 0.0402$, $n_1 = 2.617$.

$D_b = 29\,(144/0.0402)^{1/2.617} = 661.4$ mm

Bundle to shell clearance, from Fig 12.10, for a fixed tube sheet = 14 mm

So, Shell inside diameter = 661.4 + 14 = 675.4, round to 680 mm

A close baffle spacing is not needed for a condensing vapour. All that is needed is sufficient baffles to support the tubes. Take the baffle spacing as equal to the shell diameter, 680 mm.

Number of baffles = $(4 \times 10^3/ 680) - 1 = 5$

Step 13 Cost

From chapter 6, Fig 6.3, basic cost for carbon steel exchanger = £10,000

Type factor for fixed tube sheet = 0.8. Pressure factor 1.0.

So, cost = 10000 x 0.8 x 1.0 = £8000 at mid-1998 prices.

Step 14 Optimisation

The design could be improved, to make use of the full pressure drop allowance on the tube-side. If the number of tubes were reduced to, say 120, the tube-side velocity would be increased. This would increase the tube-side heat transfer, which would compensate for the smaller surface area.

The heat transfer coefficient is roughly proportional to the velocity raised to the power of 0.8.

$h_i \cong 1953\,(144/120)^{0.8} = 2344\ \mathrm{Wm^{-2}\,°C^{-1}}$, giving $U_o = 1046\ \mathrm{Wm^{-2}\,°C^{-1}}$

So the number of tubes required = $144 \times (720/1046) = 99$

Pressure drop is roughly proportional to the velocity squared.

$\Delta P_i = 0.24 \times (144/120)^2 = 0.35$ bar, still well below that allowed.

To just meet the pressure drop allowance = 0.7 - 0.03 = 0.67 bar, allowing for the drop across the nozzles, the number of tubes could be reduced to $144/(0.657/0.24)^{1/2} = 87$.

So it would be worth trying a six-pass design with 15 tubes per pass.

Solution 12.5

This is a rating problem, similar to problem 12.3. The simplest way to check if the exchanger is suitable for the thermal duty is to estimate the area required and compare it with the area available. Then check the pressure drops.

Procedure

1. Carry out a heat balance to determine the rate of heat transfer required and the water flow-rate

2. Estimate the tube-side coefficient using equation 12.15.

3. Evaluate the shell-side coefficient using Kern's method, given in section 12.9 .3.

4. Determine the overall coefficient using equation 12.2.

5. Calculate the mean temperature difference; section 12.6

6. Determine the area required, equation 12.1.

7. Calculate the surface area available
= number of tubes x (Π x tube o.d. x tube- length).

If area available exceeds that required by a sufficient margin to allow for the uncertainties in the design methods, particularly Kern's method, say +20%, accept that the exchanger will satisfy the thermal duty.

If there is not sufficient margin, more sophisticated methods should be used to check the shell-side coefficient; such a, Bell's method (using standard clearances) or a CAD method

8. Check the tube-side pressure drop, equation 12.20. Add the pressure drop over the nozzles, section 12.8.2.
.
9. Check the shell-side pressure drop, including the nozzles; use Kern's method section 12.9.3.

If the pressure drops are within limits, accept the exchanger.
If the shell-side limit is critical, a reasonable margin is needed to cover the approximate nature of the method used

Notes

1. The density of the ammonia stream will vary for the inlet to outlet due to the change in temperature. Use the mean density in the calculations.

2. The viscosity correction factor can be neglected for both streams.

Solution 12.6

First check that the critical flux will not be exceeded. Then check that the exchanger has sufficient area for the duty specified.

By interpolation, saturation temperature = 57 °C.

From steam tables, steam temp = 115.2 °C.

Duty, including sensible heat, $Q = (10{,}000/3600)(322 + 2.6(57 - 20)) = 1162$ kW

Surface area of exchanger $= (\Pi \times 30\times10^{-3} \times 4.8)50 = 22.6\ \mathrm{m}^2$

Flux $= 1162 \times 10^{3} / 22.6 = 51{,}416\ \mathrm{W/m}^2$

Critical flux, modified Zuber equation, 12.74

$$q = 0.44(45/30) \times \frac{322 \times 10^{3}}{\sqrt{(2 \mathrm{X} 50)}}(0.85 \times 9.8(535 - 14.4)14.4^{2})^{0.5} = 654{,}438\ \mathrm{W/m}^2$$

Apply the recommended factor of safety. 0.7

Critical flux for the bundle $= 0.7 \times 654438 = 458{,}107\ \mathrm{W/m}^2$

So, the operating flux will be well below the critical flux.

Use the Foster-Zuber equation, 12.62, to estimate the boiling coefficient.

Tube surface temperature = steam temperature - temperature drop across the tube wall and condensate..

$$\text{Tube wall resistance} = \frac{d_o \mathrm{Ln}\,(d_o/d_I)}{2\,k_w} = \frac{30 \times 10^{-3}\mathrm{Ln}\,(30/25)}{2 \times 50} = 0.000055\ {}^\circ\mathrm{C\ m^2W^{-1}} \qquad (12.2)$$

Take the steam coefficient as 8000 $\mathrm{Wm^{-2}\,{}^\circ C^{-1}}$; section 12.10.5.

Condensate resistance $= 1/8000 = 0.000125\ {}^\circ\mathrm{C\ m^2W^{-1}}$

Temperature drop = q x resistance $= 51416 \times (0.000055 + 0.000125) = 9.3$ °C

$T_s = 115.5 - 9.3 = 106.2\ {}^\circ\mathrm{C}$, $P_s = 17.3$ bar

$$h_{nb} = 0.00122 \left[\frac{0.094^{0.79}\,(2.6 \times 10^{3})^{0.45}\,535^{0.49}}{0.85^{0.5}\,(0.12 \times 10^{-3})^{0.29}\,(322 \times 10^{3})^{0.24} \times 14.4^{0.24}} \right]$$

$$\times\,(106.2 - 57)^{0.24}\,\{(17.3 - 6) \times 10^{5}\}^{0.75} = 4647\ \mathrm{Wm^{-2}\,{}^\circ C^{-1}} \qquad (12..62)$$

$$1/U_o = (1/5460)(30/25) + 0.000055 + 0.00125 \qquad (12..2)$$

$$U_o = 2282\ \mathrm{Wm^{-2}\,{}^\circ C^{-1}}$$

As the predominant mode of heat transfer will be pool boiling, take the driving force to be the straight difference between steam and fluid saturation temperatures.

$$\Delta T_m = 112.5 - 57 = 55.5 \text{ °C}$$

Area required = $(1163 \times 10^3)/(2282 \times 55.5) = 9.2 \text{ m}^2$ (12.1)

Area available = 22.6 m^2. So there is adequate area to fulfil the duty required; with a good margin to cover fouling and the uncertainty in the prediction of the boiling coefficient.

Solution 12.7

This a design problem, so there will be no unique solution. The solution outlined below is my first trial design . It illustrates the design procedure and methods to be used.

The physical properties of propanol were taken from Perry's Chemical Engineering Handbook and appendix D. Those for steam and water were taken from steam tables.

Propanol, heat of vaporisation = 695.2 kJ/kg , specific heat = 2.2 kJ kg^{-1} °C^{-1}.

Mass flow-rate = 30000/3600 = 8.33 kg/s

Q, condensation = 8.33 x 695.2 = 5791 kW

Q, sub-cooling = 8.33 x 22(118 – 45) = 1338 kW

For condensation, take the initial estimate of the overall coefficient as 850 Wm^{-2} °C^{-1} ; Table 12.1. For sub-cooling take the coefficient as 200 Wm^{-2} °C^{-1} , section 12.10.7.

From a heat balance, using the full temperature rise. cooling water flow-rate =

$$(5791 + 1338)/(4.2(60 – 30)) = 56.6 \text{ kg/s}$$

Temperature rise from sub-cooling = 1388/(4.2 x 56.6) = 5.8 °C

Cooling water temperature after sub-cooling = 30 + 5.8 = 35.8 °C

Condensation

118 - - - - → - - - - - 118 °C

60 - - - - - ← - - - - - - 35.8 °C

$\Delta T_M = \Delta T_{LM}$ = (118 – 60) – (118 – 35.8)]/[Ln (58/82.2)] = 69.4 °C (12.4)

Area required, A = $5791 \times 10^3/(850 \times 69.4) = 98 \text{ m}^2$

Sub-cooling

118 - - - - - → - - - - - 45 °C

35.8 - - - - ← - - - - - - 30 °C

ΔT_{LM} = [(118 – 35.8) - (45 - 30)]/[Ln (82.2/15) = 39.5 °C

R = (118 – 45)/(35.8 – 30) = 12.6, S = (35.8 – 30)/(118 – 30) = 0.07 (12.5)(12/6)

F_t = 1.0, Fig 12.19, one shell pass even tube passes. So, ΔT_M = 39.5 °C

Area required = 1338 x 10^3/(200 x 39.5) = 169 m^2

Best to use a separate sub-cooler

Condenser design

ΔT_M = ΔT_{LM} = (118 – 60) – (118 – 30)]/[Ln (58/88)] = 72 °C (12.4)

Area required = 5791 x 10^3/(850 x 72) = 95 m^2

Surface area of one tube = Π x 19 x 10^{-3} x 2.5 = 0.149 m^2

Number of tubes = 95/0.149 = 638

Put the condensing vapour in the shell.

Tube cross-sectional area = Π/4(16 x $10^{-3})^2$ = 2.01 x 10^{-4} m^2

Water velocity with one pass = (56.6/990.2)/(638 x 2.01 x 10^{-4}) = 0.45 m/s

Low, try 4 passes, 160 tubes per pass, 640 tubes

u_t = (56.6/990.2)/(160 x 2.01 x 10^{-4}) = 1.8 m/s

Looks reasonable, pressure drop should be within limit.

Outside coefficient

Use square pitch, p_t = 1.25d_o = 12.5 x 19 = 23.75 mm

Bundle diameter, D_b = 19(640 /0.158)$^{1/2.263}$ = 746 mm

Number of tubes in centre row = D_b/p_t = 746/23.75 = 32

Take N_r = 2/3 x 32 = 21

Mol mass propanol = 60.1

Density of vapour = (60.1/22.4) x (273/391) x (2.1/1) = 3.93 kg/m^3

Γ_h = W_c/LN_t = 8.33/(2.5 x 640) = 0.0052 kg/m

$$(h_c)_b = 0.95 \times 0.16 \left[\frac{740(740 - 3.93)\ 9.8}{447 \times 10^{-6} \times 0.0052}\right]^{1/3} \times 21^{-1/6} = 1207\ \mathrm{Wm^{-2}\,{}^{\circ}C^{-1}} \qquad (12.50)$$

Inside coefficient

Re = $(990.2 \times 1.8 \times 16 \times 10^{3})/(594 \times 10^{-6}) = 48010$, Pr = 3.89

From Fig 12.24, $j_h = 3.3 \times 10^{-3}$. Neglect viscosity correction

$$Nu = 3.3 \times 10^{-3} (48010)(3.89)^{0.33} = 248 \qquad (12.15\ )$$

$$h_i = 248 \times 638 \times 10^{-3}/16 \times 10^{-3} = 9889\ Wm^{-2}\ °C^{-1}$$

$$1/U_o = (1/9889)(19/16) + \frac{19 \times 10^{-3}(Ln(19/16)}{2 \times 50} + 1/1207 \qquad (12.2)$$

$U_o = 1019\ Wm^{-2}\ °C^{-1}$. Greater than the initial value, with sufficient margin to allow for fouling.

Check pressure drops

Tube-side: $u_t = 1.8$ m/s, Re = 48010, $j_f = 3.1 \times 10^{-3}$ Fig 12.24, neglect viscosity correction factor.

$$\Delta P = 4[8 \times 3.1 \times 10^{-3} (2.5/16 \times 10^{-3}) + 2.5](990.2 \times 1.8^2/2) = 62160\ N/m^2 = 62\ kN/m^2 \qquad (12.20)$$

A bit high, only 8 kN/m^2 available to for losses in nozzles.

Could try increasing the number of tubes or reducing the number of passes, or both.

Overall coefficient is tight, so could try, say, 800 tubes with two tube passes.

Shell-side: shell clearance, for split-ring floating head exchanger = 65 mm, Fig 12.10.

So, $D_s = 746 + 65 = 811$ mm

Take baffle spacing = D_s = 811, close spacing not needed for a condenser.

$$A_s = \frac{(23.75 - 19)}{23.75}(811 \times 10^{-3} \times 811 \times 10^{-3}) = 0.132\ m^2 \qquad (12.21)$$

$$u_s = (8.33/3.93)/0.132 = 16.1\ m/s$$

$$de = 12.7(23.75^2 - 0.785 \times 19^2)/19 = 18\ mm \qquad (12.22)$$

$$Re = (16.1 \times 3.93 \times 18 \times 10^{-3})/(0.01 \times 10^{-3}) = 113891$$

$j_f = 3.5 \times 10^{-2}$, Fig 12.30. Neglect viscosity correction

$\Delta P_s = 8 \times 3.5 \times 10^{-2} [(811 \times 10^{-3}/18 \times 10^{-3}) (2.5/0.811)](3.93 \times 16.1^2 /2) = 19808 \text{ N/m}^2$ (12.26)

So pressure drop based on the inlet vapour flow-rate = 19.8 kN/m^2.

This is well below the limit for the total pressure drop so there is no need to refine the estimate.

Sub-cooler design

Put propanol in shell.

$$118 \text{ - - - -}\rightarrow\text{ - - - - } 45\ °\text{C}$$

$$60 \text{ - - - -}\leftarrow\text{ - - - - } 30\ °\text{C}$$

$$\Delta T_{LM} = \frac{(118 - 60) - (45 - 30)}{\text{Ln}(58/15)} = 31.8\ °\text{C}$$

R = (118 – 45)/(60 – 30) = 2.4, S = (60 – 30)/(118 – 30) = 0.34

Correction factor F_t is indeterminant for a single shell pass exchanger, Fig 12.19.

Try two shell passes, Fig 12.20, F_t = 0.9

$\Delta T_m = 0.9 \times 31.8 = 28.6\ °\text{C}$

Could use two single shell-pass exchangers to avoid the use of a shell baffle.
I will design a two shell-pass exchanger to illustrate the method.

From table 12.1, U = 250 to 750 Wm^{-2} °C^{-1} . Try 500 Wm^{-2} °C^{-1}

$$A_s = \frac{1338 \times 10^3}{500 \times 28.6} = 94 \text{ m}^2$$

Number of tubes = 94/0.149 = 631

Tube-side coefficient

Cooling water flow-rate = 1338/(4.2 x 30) = 10.62 kg/s

$$\text{Tube side velocity, single pass} = \frac{10.62 / 990.2}{631 \times 2.56 \times 10^{-4}} = 0.066 \text{ m/s}$$

Far too low, try 8 passes, 83 tubes per pass, 664 tubes.

$u_t = \dfrac{10.62 / 990.2}{83 \times 2.56 \times 10^{-4}} = 0.505$ m/s

$Re = \dfrac{990.2 \times 0.505 \times 16 \times 10^{-3}}{594 \times 10^{-6}} = 13469$

$j_h \cong 4.0 \times 10^{-3}$, Fig 12.23

$Nu = 4.0 \times 10^{-3} \times 13469 \times (3.89)^{0.33} = 84.3$

$h_i = 84.3 \times (638 \times 10^{-3} / 16 \times 10^{-3}) = 3361$ $Wm^{-2}\,°C^{-1}$

Shell-side coefficient

$D_b = 19(664/0.0365)^{1/2.675} = 743$ mm

Use 25% cut baffles, spacing 0.5 D_s = 372 mm.

Triangular pitch, $p_t = 1.25\, d_o$

$$A_s = \frac{23.75 - 19}{23.75}(743 \times 10^{-3} \times 372 \times 10^{-3}) = 0.055 \text{ m}^2 \qquad (12.21)$$

For two shell passes, the cross-flow area is taken as half that given by equation 12.21, as the shell baffle divides the shell cross-section into two equal halves.

So, $u_s = (8.33/752) / (0.055/2) = 0.403$ m/s

$d_e = (1.10/19)/(23.75^2 - 0.917 \times 19^2) = 13.5$ mm

$Re = \dfrac{752 \times 0.403 \times 13.5 \times 10^{-3}}{508 \times 10^{-6}} = 8054$

From Fig 12.29, $j_h = 6.3 \times 10^{-3}$. Neglect viscosity correction

$Pr = (2.2 \times 10^3 \times 508 \times 10^{-6}/ 0.164) = 6.2$

$Nu = 6.3 \times 10^{-3} \times 8054 \times (6.2)^{0.33} = 92.6$

$h_s = 92.6 \times (0.164/13.5 \times 10^{-3}) = 1125$ $Wm^{-2}\,°C^{-1}$

$$1/U_o = (1/3316)(19/16) + \frac{19 \times 10^{-3} \text{ Ln } (19/16)}{2 \times 50} + 1/1125$$

$U_o = 781$ $Wm^{-2}\,°C^{-1}$ well above the trial value of 500 $Wm^{-2}\,°C^{-1}$.

Reasonable margin to allow for fouling; accept design but check pressure drops.

Tube-side

For Re = 13469, j_f = 4.5×10^{-3}, Fig 12.24

$\Delta P_t = 8\,[8 \times 4.5 \times 10^{-3}\,(2.5/16 \times 10^{-3}) + 2.5)]\,(990.2 \times 0.505^2)/2 = 8207\ \text{N/m}^2$

Well below the limit set for the cooling water.

Shell-side

From Fig 12,10 clearance for split-ring floating head exchanger = 65 mm
D_s = 743 + 65 = 808 mm

For Re = 8054, j_f = 5.0×10^{-2} , Fig 12.30

For a two pass-shell, the number of tube crosses will be double that given by the term L/l_b in equation 12.26. There will be set of cross-baffles above the shell baffle and a set below, which doubles the path length.

So, $\Delta P_s = 8 \times 5.0 \times 10^{-2}\,(808 \times 10^{-3} / 12 \times 10^{-3}) \times 2(2.5/0.372)\,(752 \times 0.403^2)/2$

$= 19650\ \text{N/m}^2 = 19.7\ \text{kN/m}^2$

Looks reasonable. The condensate would be pumped through the sub-cooler.

Solution 12.8

The design method will follow that used in problem 12.6, except that the condensing coefficient will be estimated for vertical tubes; section 12.10.3.

Put he condensate in the shell.

The condensing coefficient will be lower for vertical tubes, so the number of tubes will need to be increased. It would be better to increase the tube length to obtain the increased area required but the tube length is fixed.

The sub-cooler design will be the same as that determined in the solution to problem 12.7.

Solution 12.9

The design procedure will follow that illustrated in the solution to 12.7.

As the vapour is only partially condensed, from a non-condensable gas, the approximate methods given in section 12.10.8 need to be used to estimate the condensing coefficient.

Solution 12.10

As the process fluid is a pure liquid, Frank and Pricket's method can be used to give a conservative estimate of the number of tubes required. See example 12.9.

Solution 12.11

This a design problem, so there will be no unique solution. The solution outlined below is my first trial design . It illustrates the design procedure and methods to be used.

Mass flow-rate = 10000/3600 = 2.78 kg/s

Duty = 2.78 [0.99(10 – 10) + 260] = 722.8 kW

The water outlet temperature is not fixed. The most economic flow will depend on how the water is heated. The simplest method would be by the injection of live steam. The heated water would be recirculated. As a trial value, take the water outlet temperature as 40 °C.

Water flow-rate = 722.8/[4.18(50 – 40)] = 17.3 kg/s

10 °C - - - - - →- - - - - - 10 °C

50 °C - - - - - -← - - - - - 40 °C

$\Delta T_M = \Delta T_{LM} = (40 - 30)/\mathrm{Ln}(40/30) = 34.8$ °C (12.4)

The coefficient for vaporisation will be high, around 5000 $\mathrm{Wm^{-2}\,{}^{\circ}C^{-1}}$. That for the hot water will be lower, around 2000 $\mathrm{Wm^{-2}\,{}^{\circ}C^{-1}}$. So, take the overall coefficient as 1500 $\mathrm{Wm^{-2}\,{}^{\circ}C^{-1}}$.

Area required = $(722.8 \times 10^3)/(1500 \times 34.8) = 13.8\ \mathrm{m}^2$ (12.1)

Surface area of one U-tube = $\Pi \times 25 \times 10^{-3} \times 6 = 0.47\ \mathrm{m}^2$

Number of U-tubes required = 12.8/0.47 = 30

Shell-side coefficient

Heat flux, q = $722.8 \times 10^3/(30 \times 0.47) = 51262\ \mathrm{W/m^2}$

Taking k_w for stainless steel = 16 $\mathrm{W\,m^{-1}\ {}^{\circ}C^{-1}}$

Resistance of tube wall, $R = \dfrac{25 \times 10^{-3}\mathrm{Ln}(25/21)}{2 \quad \times 16} = 0.000136\ (\mathrm{Wm^{-2}\,{}^{\circ}C^{-1}})^{-1}$

ΔT cross tube wall = q x R = 51262 x 0.000136 = 7 °C

So mean tube wall surface temperature, T_w = 45 - 7 = 38 °C

$\mathrm{Ln}(P_w) = 9.34 - 1978/(38 + 246)$, $P_w = 10.75$ bar

$$h_{nb} = 0.0012\left[\frac{0.13^{0.79} \times 990^{0.45} \times 1440^{0.49}}{0.013^{0.5}\,(0.3 \times 10^{-3})^{0.29}\,(260 \times 10^{3})^{0.24}\,16.4^{0.24}}\right](38 - 10)^{0.24}\,[(10.75 - 5)10^{5}]^{0.75}$$

$$= 21043 \text{ Wm}^{-2}\,°\text{C}^{-1} \qquad (12.62)$$

Tube side coefficient

Properties of water at 45 °C, from steam tables: $\rho = 990.2$ kg/m^3,
$\mu = 594 \times 10^{-6}$ N m^{-2} s, $k = 638 \times 10^{-3}$ W m^{-1} °C^{-1}, $C_p = 4.18$ kJ kg^{-1}°C^{-1}, Pr = 3.89

Cross-sectional area of one tube = $\Pi/4 \times (21 \times 10^{-3})^2 = 3.46 \times 10^{-4}$

$u_t = (17.3/990.2)/(30 \times 3.46 \times 10^{-4}) = 1.68$ m/s

$Re = 990.2 \times 1.68 \times 21 \times 10^{-3})/594 \times 10^{-6} = 58{,}812$

$j_h = 3.2 \times 10^{-3}$, Fig 12.23. Neglect the viscosity correction

$$Nu = 3.2 \times 10^{-3} \times 58812 \times 3.89^{0.33} = 294.6 \qquad (12.15)$$

$h_i = 294.6 \times (638 \times 10^{-3}/21 \times 10^{-3}) = 8950$ Wm^{-2} °C^{-1}

$1/U = 1/21043 + 0.000136 + 1/8950$

$U = 3387$ Wm^{-2} °C^{-1} Well above the assumed value.

Check maximum heat flux

Take the tube pitch to be 1.5 x tube o.d., on a square pitch, to allow for vapour flow.

$p_t = 25 \times 1.5 = 37.5$ mm

$N_t = 30 \times 2 = 60$ (U-tubes)

$$q_{ch} = 0.44 \times (37.5/25)(260 \times 10^3)[0.013 \times 9.8(1440 - 16.3)16.3^2]^{0.25}$$

$$= 2{,}542{,}483 \text{ W/m}^2 \qquad (12.74)$$

Apply a 0.7 factor of safety, = 1,779,738 W/m^2

Actual flux = 51,262 W/m^2, well below the maximum.

Check tube-side pressure drop

For $Re = 58{,}812$, from Fig 12.24, $j_f = 3.2 \times 10^{-3}$

L in equation 12.20 = half U-tube length = 3m

$\Delta P_t = 2[8 \times 3.2 \times 10^{-3} (3/21 \times 10^{-3}) = 2.5] 990.2 \times 1.68^2/2$

$=17208$ N/m^2 = 0.17 bar, well within the limit specified

Shell design

A shell similar to that designed in example 12.11 could be used. Or, the bundle could be inserted in a simple, vertical, pressure vessel, with sufficient height to provide adequate disengagment of the liquid drops; see section 10.9.2.

Solution 12.12

The properties of the solutions to be taken as for water. As there is little difference in the mean temperatures of the two streams, use the properties at 45 °C.
From steam tables: $\rho = 990.2$ kg/m^3, $\mu = 594 \times 10^{-6}$ N m^{-2} s, $k = 638 \times 10^{-3}$ W m^{-1} °C^{-1}, Cp = 4.18 kJ kg^{-1}°C^{-1}, Pr = 3.89.

The temperature change of the cooling water is the same as that of the solution, so the flow-rates will be the same.

Flow-rate = 200000/3600 = 55.6 kg/s

There are 329 plates which gives 329 – 1 flow channels.

The flow arrangement is 2:2, giving 4 passes

So, the number of channels per pass = (329 – 1)/4 = 82

Cross-sectional area of a channel = $0.5 \times 3 \times 10^{-3} = 1.5 \times 10^{-3}$ m^2

The velocity through a channel = $(55.6/990.2)/(82 \times 1.5 \times 10^{-3}) = 0.46$ m/s

Equivalent diameter, $d_e = 2 \times 3 = 6$ mm

$Re = (990.2 \times 0.46 \times 6 \times 10^{-3})/594 \times 10^{-6} = 4601$

$Nu = 0.26 (4601)^{0.65} \times (3.89)^{0.4} = 107.6$ (12.77)

Neglecting the viscosity correction factor

$h_p = 107.6 \times (638 \times 10^{-3} / 6 \times 10^{-3}) = 11441$ Wm^{-2} °C^{-1}

As the flow-rates and physical properties are the same for both streams the coefficients can be taken as the same.

The plate material is not given, stainless steel would be suitable and as it has a relatively low thermal conductivity will give a conservative estimate of the overall coefficient.

Take thermal conductivity of plate = 16 W m^{-1} $°C^{-1}$

$1/U = 1/11441 + 0.75 \times 10^{-3}/16 + 1/11441$

$U = 4511$ Wm^{-2} $°C^{-1}$

$$70 - - - \rightarrow - - -\ 30\ °C$$

$$60 - - - \leftarrow - - -\ 20\ °C$$

As the terminal temperature differences are the same, $\Delta T_{LM} = \Delta T = 10$ °C

$NTU = (70 - 30)/10 = 4$

F_t from Fig 12.62 = 0.87

$\Delta T_M = 10 \times 0.87 = 8.7$ °C

Duty, $Q = 55.6 \times 4.18(70 - 30) = 9296.3$ kW

Area required = $(9296.3 \times 10^3)/(4511 \times 8.7) = 236.9$ m^2

Number of thermal plates = total – 2 end plates = 329 – 2 = 327

Area available = $327(1.5 \times 0.5) = 245$ m^2

So exchanger should be satisfactory. but there is little margin for fouling.

Pressure drop

The pressure drop will be the same for each stream

$j_f = 0.6 \times (4601)^{-0.3} = 4.8 \times 10^{-2}$

Lp, two passes = $2 \times 1.5 = 3$ m

$\Delta P_p = 8 \times 4.8 \times 10^{-2} (3/6\times 10^{-3})\, 990.2 \times 0.46^2/2 = 20115\ N/m^2$ (12.78)

Port area = $\Pi \times (0.15^2)/4 = 17.7 \times 10^{-3}\ m^2$

Velocity $u_{pt} = (55.6/990.2)/(17.7 \times 10^{-3}) = 3.17$ m/s

$\Delta P_{pt} = 1.3 \times 990.2 \times (3.17^2/2) \times 2 = 12936\ N/m^2$ (12.79)

Total pressure drop for each stream = 20115 + 12936 = 33052 N/m^2

= 0.33 bar