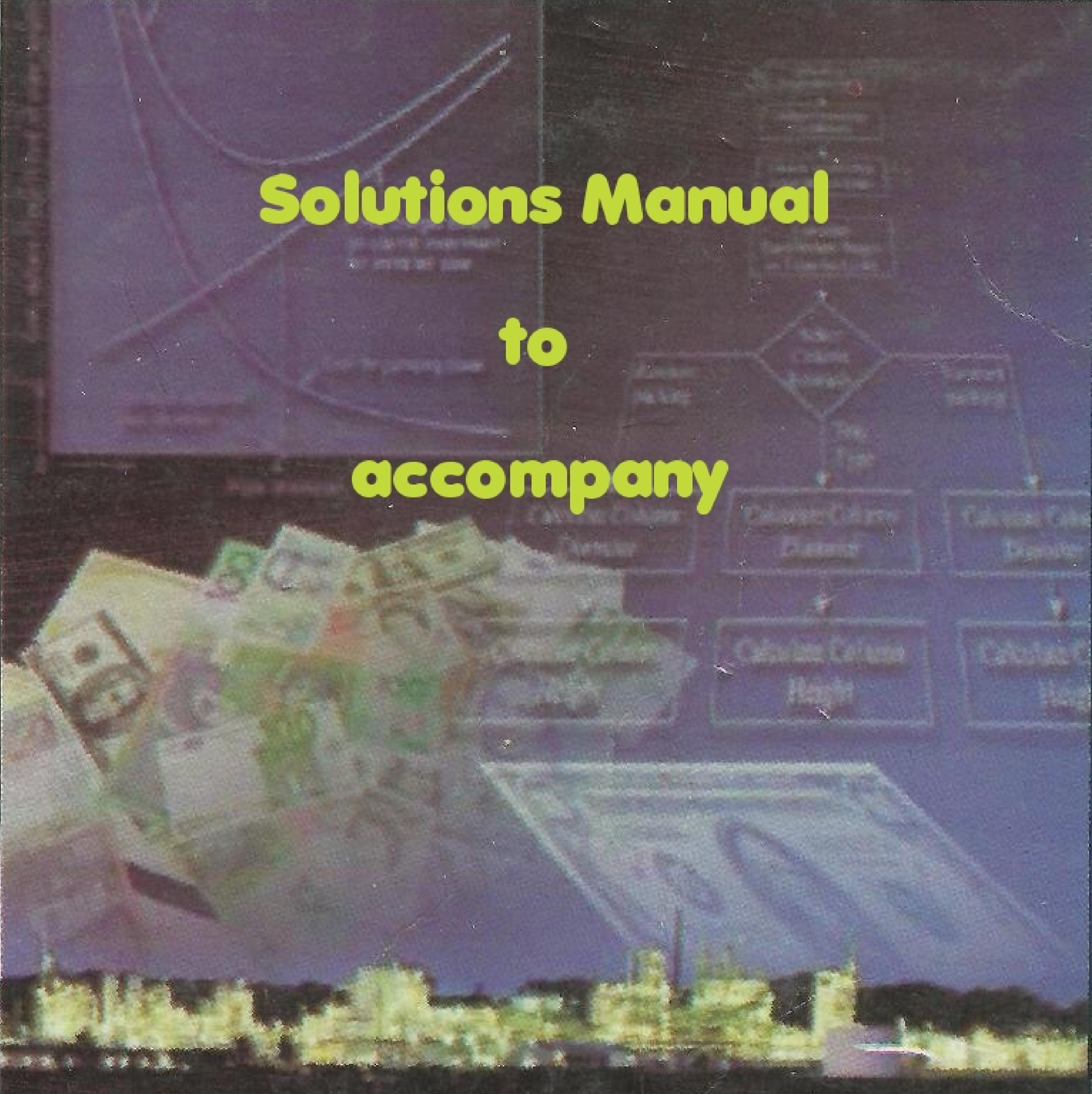

Solutions Manual

to

accompany

Plant Design and Economics for Chemical Engineers

Fifth Edition

Max S. Peters Klaus D. Timmerhaus Ronald E. West

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McGraw-Hill Science

2-1

The vapor pressure of mercury is obtained from <u>Perry's Chemical Engineers' Handbook</u>, 6th ed., p. 3-201 as

T, °C	Vapor press, bar
10	7.101×10^{-7}
20	1.729×10^{-6}
30	3.968×10^{-6}

By interpolation, the vapor pressure of mercury at 23°C is 2.2×10^{-6} bar. For low pressures, at equilibrium, the vapor phase mol fraction of mercury can be obtained with the use of Dalton's law, or

$y = p_v^\circ / p$ where p = system pressure of 1.013 bar

$y = 2.2 \times 10^{-6} / 1.013 = 2.17 \times 10^{-6}$ kg mols Hg / kg mols air

but $y =$ mols Hg / (mols air + mols Hg) $\cong$ mols Hg / mols air

The specific volume of air is obtained from the ideal gas relation $v = RT/p = (8314)(296.1)/101{,}300 = 24.305\ \mathrm{m^3/kg\ mol}$

The concentration of mercury in air at equilibrium is then

$$c = (2.17 \times 10^{-6} / 24.305)(200.6) = 17.9 \times 10^{-6}\ \mathrm{kg\ Hg/m^3}$$
$$= 17.9\ \mathrm{mg\ Hg/m^3}$$

Since the PEL (permissible exposure limit is $0.1\ \mathrm{mg\ Hg/m^3}$, the concentration of mercury in the air greatly exceeds this limit and therefore is <u>not acceptable.</u>

Answer

2-2

This problem has been simplified considerably with a number of assumptions. Actually, the corrosion rate for the iron (carbon steel) over a year is not in a fixed oxygen level atmosphere. It is most likely that the corrosion rate of iron is limited by the diffusion rate of oxygen to the tank surface. Under the circumstances, the analysis would require a corrosion model, information on the initial rate before corrosion product buildup, and need some indication of the diffusivity of the oxygen in the corrosion layer. As indicated in the problem, the reaction rate is a function of the surface are inside the tank. For a 7.6-m (inside) diameter spherical container, the inside area is given by

$$A = 4\pi r_i^2 = \pi D_i^2 = \pi (7.6)^2 = 181.5 \text{ m}^2$$

Iron corrosion $= (181.5)(0.127 \times 10^{-3})(7850)$ $\qquad \rho_{Fe} = 7850 \text{ kg/m}^3$

$= 180.9$ kg/y

$= (180.9)/(365)(24) = 0.02065$ kg/h

$= 0.02065/55 = 3.7545 \times 10^{-4}$ kg mol/h

Oxygen rate $= (3.7545 \times 10^{-4})(1.5 \text{ kg mol } O_2 / 2.0 \text{ kg mol Fe})$

$= 2.816 \times 10^{-4}$ kg mol/h

Volume of sphere $= (4/3)\pi r_i^3 = \pi D_i^3/6 = \pi (7.6)^3/6 = 229.8 \text{ m}^3$

At room temperature (21.1°C) and 1 atm, the specific volume inside the tank, assuming an ideal gas, is

$$v = (22.4)(294.2/273.1) = 24.147 \text{ m}^3/\text{kg mol air}$$

The oxygen concentration under these conditions is $(0.21)(24.147) = 8.696$ kg mol O_2/m^3

For a first order reaction,

$\frac{dc}{d\theta} = -kc$ where k = reaction rate constant, θ = time, and c = concentration

2-2 (continued)

When $c = 8.696 \times 10^{-4}$ kg mol/m^3, the O_2 rate is 2.816×10^{-4} kgmol/h, and the tank volume is 229.8 m^3. At this condition

$$\frac{dc}{d\theta} = \frac{-2.816 \times 10^{-4}}{229.8} = -1.225 \times 10^{-6} \text{ kgmol/m}^3 \cdot \text{h}$$

Solving for the reaction rate constant gives

$$k = -\left(\frac{dc}{d\theta}\right)/c = (1.225 \times 10^{-6})/8.696 \times 10^{-3} = 1.408 \times 10^{-4} \text{ h}$$

Now solve for the time when O_2 concentration is reduced to 19.5 mol percent. Integrating the original rate equation results in

$$\ln(c_1/c_2) = k(\theta_2 - \theta_1) \qquad \text{where } \theta_1 = 0$$

$$\theta_2 = \ln(c_1/c_2)/k$$

$$= \ln(0.21/0.195)/1.408 \times 10^{-4}$$

$$= \underline{\underline{526 \text{ h or approx. } 22 \text{ days}}}$$

Answer

2-3

Assume that the dilution air is available at a temperature of 20°C and a pressure of 1 atm. The evaporation rate is 7.5 g/min or 0.0075 kg/min; Thus,

$\dot{n}_{vc} = 0.0075/78 = 9.62 \times 10^{-5}$ kg mol vinyl chloride/min

At 1 atm use the ideal gas to obtain the volumetric flow rate

$\dot{m}_{v,vc}\ n_{cv}RT/p = (9.62 \times 10^{-5})(8314)(293/1.013 \times 10^{5}$

$= 2.31 \times 10^{-3}\ m^3/min$

Since the PEL for vinyl chloride is 1ppm,

$\dot{m}_{v,air} = (\dot{m}_{v,vc})(10^{6}) = (2.31 \times 10^{-3})(10^{6})$

$= 2.31 \times 10^{3}\ m^3/min$

If a safety factor of 5 is required,

$\dot{m}_{v,air} = (5)(2.31 \times 10^{3})/60 = \underline{\underline{192.5\ m^3/s}}$

Answer

This is a fairly high flow rate and would need one or two large blowers.

2-4

The density of benzene at 22°C is 879 kg/m^3. The molecular weight is 78.1 kg/kg mol.

Since the pressure is low, assume the ideal gas law is applicable to the air stream that is used for ventilation.

$$\dot{n}_{Bz} = \left(2.0 \times 10^{-6} \frac{m^3}{min}\right)\left(879 \frac{kg}{m^3}\right) / 78.1 \frac{kg}{kg\,mol}$$

$$= 2.2506 \times 10^{-5} \text{ kg mol/min}$$

$$\dot{m}_{v,Bz}\ n RT/p = \left(2.2506 \times 10^{-5} \frac{kg\ mol}{min}\right)\left(8314 \frac{J}{kg\,mol \cdot K}\right)(295\ K)$$

$$6.27 \times 10^{-4}\ m^3/min$$

Assume complete mixing of the benzene with the dilution air

$$\dot{m}_{v,air} = \left(6.27 \times 10^{-4} \frac{m^3\ Bz}{min}\right)\left(10^6 \frac{m^3\,air/min}{m^3\,Bz/min}\right)$$

$$= \underline{\underline{627\ m^3\ air/min}}$$

Answer

2-5

To determine the length of time the cannister will be usable, we will need to find a relationship between the amount adsorbed and $(T/V)\log(f_s/f)$. Let $x = (T/V)\log(f_s/f)$ and y the amount adsorbed in $cm^3(liq)/100\ g$ charcoal. From the data given,

when $y = 1 \quad x = 21$
$y = 10 \quad x = 11$

assuming a linear relation of $\log y = ax + b$

$\log 1 = 21a + b$
$\log 10 = 11a + b$
or $\log(1/10) = 10a \quad a = -0.1$
$b = \log 10 - 11(-0.1) \quad b = 2.1$

The molecular weight of dichloropropane (DCP) is 113, density is $1160\ kg/m^3$, and molar volume $= 0.1\ m^3/kg\ mol$. The vapor pressure from Perry's Chemical Engineers' Handbook, 6 ed., p.354,

Temp., °C	Vapor press., mm Hg
19.4	40
28	60

Assuming the Antoine equation can be used to represent the vapor pressure of DCP

$\log p = A + B/T$
$\log 40 = A + B/292.55$
$\log 60 = A + B/301.15$

Solving the latter two equations simutaneously results in

$B = -1802 \quad A = 7.76$

At 26.9°C or 300.05 K

$\log p = 7.76 + (-1902/300.05) = 1.755$
$p = 56.9\ mm\ Hg$

Determine the fugacity of DCP by assuming that it is essentially equal to the partial pressure. Thus,

2-5 (continued)

$f \simeq p$

$p = y p_T$ where p_T = total pressure and $y = 750/10^6$

$p = (750/10^6)(760) = 0.57 \text{ mm Hg}$

therefore, $f_s/f = 56.9/0.57 = 99.82$

$x = (T/V) \log (f_s/f) = (300/100) \log (99.82) = 5.99$

Now determine how much gas has been adsorbed

$\log y = -0.1x + 2.1 = (-0.1)(5.99) + 2.1 = 1.503$

$y = 31.84 \text{ cm}^3/100 \text{ g}$ charcoal

For 100% saturation, amount adsorbed per 100g charcoal is

$m = (31.84 \text{ cm}^3/100 \text{ g C})(1.16 \text{ g/cm}^3) = 36.93 \text{ g}$

At 82% saturation, amount adsorbed before breakthrough is

$m_{82\%} = (36.93)(0.82) = 30.3 \text{ g}$

Now determine volume of air associated with 30.3 g DCP

$$n_{air} = \left(\frac{30.3 \text{ g DCP}}{113 \frac{\text{g DCP}}{\text{g mol DCP}}} \right) \left(\frac{10^6 \text{ g mol air}}{750 \text{ g mol DCP}} \right) = 357.5 \text{ g mol air}$$

$$\text{Vol of air} = nRT/p = \frac{(357.5 \text{ g mol})}{1 \text{ atm}} \left(82.06 \frac{\text{atm} \cdot \text{cm}^3}{\text{g mol} \cdot \text{K}} \right) (300 \text{ K})$$

$$= 8.8 \times 10^6 \text{ cm}^3 = 8{,}800 \text{ ℓ} = 8.8 \text{ m}^3$$

At 45 ℓ/min

$$\theta = (8{,}800 \text{ ℓ})/(45 \text{ ℓ/min}) = \underline{\underline{195 \text{ min} = 3.25 \text{ h}}}$$

Answer

Cannister should last 3.25h before break-through occurs.

2-6

If the mixture is at the flash point

$$\sum \frac{y_i}{LFL_i} = 1.0$$

Assuming Raoult's law applies

$$y_i = p_{v,i}\, x_i / p$$

where $p_{v,i}$ is the vapor pressure of component i, p the pressure, and x_i and y_i the mol fraction of component in the liquid and vapor, respectively. Vapor pressure of C_8 and C_9 are available from <u>Perry's Chemical Engineers' Handbook</u>, 6th ed., p. 3-58

Assume flash temperature of 18°C

Comp	x_i	$p_{v,i}$, atm	y_i	LFL_i
C_8	0.5	0.0123	0.00615	0.01
C_9	0.5	0.00365	0.00183	0.008

$$\sum \frac{y_i}{LFL_i} = \frac{0.00615}{0.01} + \frac{0.00183}{0.008} = 0.844 \text{ temp. too low}$$

Assume flash temperature of 21.1°C

Comp	x_i	$p_{v,i}$, atm	y_i	LFL_i
C_8	0.5	0.0145	0.00725	0.01
C_9	0.5	0.0044	0.0022	0.008

$$\sum \frac{y_i}{LFL_i} = \frac{0.00725}{0.01} + \frac{0.0022}{0.008} \cong 1.0$$

The flash point for this mixture is <u>21.1°C</u>
Answer

2-7

At equilibrium, $f_{i,L} = f_{i,v}$

Since

$$f_{i,L} = \gamma_i f_{o,L} x_i$$

$$f_{i,v} = \varphi_i f_{o,v} y_i$$

$$\gamma_i f_{o,L} x_i = \varphi_i f_{o,v} y_i$$

At low pressures

$$f_{o,v} = p$$

$$f_{o,L} = p_v^{\circ}$$

$$\varphi_i = 1.0$$

Therefore $\gamma_i p_v^{\circ} x_i = p y_i$

For an ideal solution, $\gamma_i = 1$

and $x_i p_v^{\circ} = p y_i$

Definitions

$f_{i,L}$ = fugacity of liquid

$f_{i,v}$ = fugacity of vapor

γ_i = liquid phase activity coeff,

φ_i = vapor phase activity coeff.

$f_{o,L}$ = pure comp. liq. fugacity

$f_{o,v}$ = pure comp. vapor fugacity

p = total pressure

p_v° = vapor pressure

If the acetone has no air dissolved in it, $x_i = 1.0$ and y_i is the LFL or $y_i = 0.0255$.

At a total pressure of 1 atm,

$$p_v^{\circ} = y_i p / x_i$$

$$= (0.0255)(1.0)/1.0$$

$$= 0.0255 \text{ atm} = 19.38 \text{ mm Hg}$$

Acetone has a vapor pressure of 19.38 mm Hg at a temperature of -21.4 °C (<u>Perry's Chemical Engineers' Handbook</u>, 6th ed., p. 3-50.

$$T_{flash} \text{ (calculated)} = \underline{\underline{-21.4\,°C}}$$

Answer

The handbook <u>Dangerous Properties of Industrial Materials</u> provides a measured flash point for acetone of -17.8°C.

2-8

At equilibrium, acetone, benzene, ethyl ether, and n-pentane are too rich to ignite. Carbon disulfide and methyl alcohol will always be in the flammable range. It should be noted that all of the vaporizing liquids can be ignited if not at equilibrium. Some are also quite toxic.

Answer

The types of fire extinguishers recommended when a small spill is ignited are as follows:

Liquid	Extinguishers recommended
acetone	alcohol foam, dry chemical, CO_2
benzene	foam, dry chemical, CO_2
carbon disulfide	foam, water blanket
ethyl ether	alcohol foam, dry chemical, CO_2
methyl alcohol	alcohol foam, dry chemical, CO_2
n-pentane	foam, dry chemical, CO_2

Answer

2-9

The stoichiometric balance for the reaction is given by

$$C_3H_8 + 5O_2 + 5(3.76)N_2 \longrightarrow 3CO_2 + 4H_2O + 5(3.76)N_2$$

For 200% theoretical air, the balance becomes

$$C_3H_8 + 10O_2 + 10(3.76)N_2 \longrightarrow 3CO_2 + 4H_2O + 5O_2 + 10(3.76)N_2$$

Since all the energy generated from the combustion process is absorbed by the product gases, the final temperature of the products is determined by means of an adiabatic flame temperature calculation. The heat capacities for each component is a function of the temperature and requires a knowledge of the Cp-temperature relation of those components. Present software is available and can be used to rapidly converge with a heat balance on the appropriate flame temperature. When this is done, a temperature of $\underline{\underline{1527\,K}}$ is obtained.

Answer

The iterative calculation can also be done by using an enthalpy table as a function of temperature for each of the coponents. The latter is available in Holman's Thermodynamics, 4th ed., Table A-18, pp 755-6 and is used in the hand calculation of this problem. The energy balance is

$$(\bar{h}_f^\circ)_{C_3H_8} = 3(\bar{h}_f + \bar{h}_T - \bar{h}_{298})_{CO_2} + 4(\bar{h}_f^\circ + \bar{h}_T - \bar{h}_{298})_{H_2O} + 5(\bar{h}_f^\circ + \bar{h}_T - \bar{h}_{298})_{O_2}$$
$$+ 37.6(\bar{h}_f + \bar{h}_T - \bar{h}_{298})_{N_2}$$

where $\bar{h}_f^\circ$ is the molar heat of formation, $\bar{h}_T$ the molar enthalpy at temperature T, and $\bar{h}_{298}$ the molar enthalpy at 298 K for the components as designated by the subscripts. Molar heats of formation for these components are from Table A-6 of the same text.

Heat losses to the surroundings are assumed to be negligible.

2-9 (continued)

To obtain the final temperature, assume different values of temperature until the energy balance is consistent. Units in kJ/kg mol.

Assume T = 1555 K

$$-103{,}925 = 3(-393{,}782 + 74{,}392 - 9374) + 4(-241{,}997 + 60{,}680 - 9904)$$
$$+ 5(51{,}307 - 8664) + 37.6(49{,}034 - 8676)$$

$$-103{,}925 \neq -20{,}500$$

Temperature assumed is too high; assume T = 1500 K

$$-103{,}925 = 3(-393{,}782 + 71{,}131 - 9374) + 4(-241{,}997 + 58{,}050 - 9904)$$
$$+ 5(49{,}272 - 8664) + 37.6(47{,}092 - 8676)$$

$$-103{,}925 \neq -123{,}997$$

Temperature assumed is too low. By interpolation, the final temperature under adiabatic conditions is 1511 K

The ideal gas law is valid at low pressures and high temperatures. Thus,

$$V_f = n_f RT/p$$

$$n_f = 3 + 4 + 5 + 37.6 = 49.6 \text{ kg mol/kg mol fuel}$$
$$= 4960 \text{ kg mol}/100 \text{ kg mol fuel}$$

$$V_f = (4960)(8314)(1511)/101{,}300 = 615{,}000 \text{ m}^3$$
$$= \underline{\underline{6.15 \times 10^5 \text{ m}^3}}$$

Answer

Ratio of final volume to initial volume is given by

$$V_f/V_i = \frac{n_f T_f}{n_i T_i} = \frac{(4960)(1511)}{(4860)(298)} = 5.17$$

2-10

Since the lower flammability limit (LFL) for acetone vapor in air is 2.55%, the maximum vapor pressure the acetone could have would be

$$p_v^\circ = (0.0255)(760) = 19.38 \text{ mm Hg}$$

The temperature equivalent to this vapor pressure may be determined utilizing the experimental data provided in Perry's Chemical Engineers' Handbook, 6th ed., p. 3-50 and the use of the Antoine equation with the form

$$\ln p_v^\circ = a - b/(t + 273)$$

The vapor pressure of acetone in the range of interest is

t, °C	p_v°, mm Hg
-40.5	5
-20.8	20

Substituting these values in the expression

$$\ln 5 = a - b/(-40.5 + 273)$$
$$\ln 20 = a - b/(-20.8 + 273)$$

Solving the equations simultaneously gives $a = 19.36$ and $b = 4126.3$. Thus,

$$\ln 19.38 = 19.36 - 4126.3/(t + 273)$$
$$t = -21.3°\text{C}$$

The acetone would have to be cooled below -21.3°C to avoid an acetone-air mixture that could be ignited with a discharge of static electricity. Such cooling is not very practical and the use of an inert gas blanket should be considered during the filling operation.

Answer

2-11

The decomposition reaction is the following

$$C_2H_2 \rightleftharpoons 2C + H_2$$

Component	mols in	mols at temp. T	mol fraction
C_2H_2	1	$1-x$	$1-x$
C	0	$2x$	0*
H_2	0	x	x

*Assumes solid carbon product

Using the nomenclature and methodology from Sandler's <u>Chemical and Engineering Thermodynamics</u>

$$K_a(T) = \frac{a_C^2 a_{H_2}}{a_{C_2H_2}} \simeq \frac{y_{H_2}}{y_{C_2H_2}} = \frac{x}{1-x}$$ Where x = extent of reac.

$$K_a(T) = e^{-(\Delta G_{rx}(T)/RT)}$$

Assume for both carbon and hydrogen

$$\Delta G_f(25°C) = 0 \quad \text{and} \quad \Delta H_f(25°C) = 0$$

$$\Delta G_{rx}(25°C) = -\Delta G_f(25°C)_{C_2H_2} = -50{,}000 \text{ cal/g mol}$$

$$\Delta H_{rx}(25°C) = -\Delta H_f(25°C)_{C_2H_2} = -54{,}194 \text{ cal/g mol}$$

The heat capacities for C, H_2 and C_2H_2 fit the relation $C_{p_i} = a_i + b_iT + c_iT^2 + d_iT^3 + e_i/T^2$ cal/g mol·K where T is in kelvins.

Comp.	a	$b \times 10^2$	$c \times 10^5$	$d \times 10^9$	e
C	2.673	0.2617	–	–	116,900
H_2	6.952	−0.04576	0.09563	−0.2079	–
C_2H_2	5.21	2.2008	−1.559	4.349	–

Data are from Sadler and <u>Perry's Chemical Engineers' Handbook</u>

For heats of reaction at other temperatures besides 25°C or T_1

$$\Delta H_{rx}^\circ(T) = \Delta H_{rx}^\circ(T_1) + \Delta a(T-T_1) + (\Delta b/2)(T^2-T_1^2) + (\Delta c/3)(T^3-T_1^3) + (\Delta d/4)(T^4-T_1^4) - \Delta e\left(\frac{1}{T}-\frac{1}{T_1}\right)$$

Where $\Delta a = \sum \nu_i a_i$, etc. and ν_i = stoichiometric coefficient

2-11 (continued, 1)

$K_a(T)$ for other temperatures besides at T_1 or 25 °C

$$\frac{\ln K_a(T)}{\ln K_a(T_1)} = \frac{\Delta a}{R}\ln\frac{T}{T_1} + \frac{\Delta b}{2R}(T-T_1) + \frac{\Delta c}{6R}(T^2-T_1^2) + \frac{\Delta d}{12R}(T^3-T_1^3) + \frac{\Delta e}{2}\left(\frac{1}{T^2}-\frac{1}{T_1^2}\right) + \frac{1}{R}\left\{-\Delta H^\circ_{rx}(T_1) + \Delta a T_1 + \frac{\Delta b}{2}T_1^2 + \frac{\Delta c}{3}T_1^3 + \frac{\Delta d}{4}T_1^4 - \frac{\Delta e}{T_1}\right\}\left(\frac{1}{T}-\frac{1}{T_1}\right) \qquad (1)$$

For this reaction $K_a(T) = X/(1-X)$ or

$$X = \frac{K_a(T)}{1+K_a(T)} \qquad (2)$$

From an energy balance for an adiabatic constant volume reaction

$$X = \frac{-\Sigma\int m_{i,in}\, C_{v_i}\, dT}{\Delta H^\circ_{rx}(T)} \qquad (3)$$

where $n_{i,in}$ are the mols that enter into the reaction. Using the heat capacity constants tabulated earlier

$\Delta a = 7.088$

$\Delta b = -1.7232 \times 10^{-2}$

$\Delta c = 1.6546 \times 10^{-5}$

$\Delta d = -4.5569 \times 10^{-9}$

$\Delta e = 2.338 \times 10^{5}$

$$K_a(25\,°C) = e^{-[(-50{,}000)/(1.987)(298.1)]} = 4.5 \times 10^{36}$$

Since the final temperature will be relatively high, the ideal gas law may be assumed to be valid. Thus,

$$C_{v,i} = C_{p,i} - R = a_i - R + b_i T + c_i T^2 + d_i T^3 + e_i/T^2$$

$\Sigma[n_i(a_i - R)]_{in} = 3.223$ $\Sigma(n_i b_i)_{in} = 2.2008 \times 10^{-2}$ $\Sigma(n_i c_i)_{in} = -1.559 \times 10^{-5}$

$\Sigma(n_i d_i)_{in} = 9.34 \times 10^{-9}$ $\Sigma(n_i e_i)_{in} = 0$

With this information, Eqs. (1), (2) and (3) can be solved simultaneously. The results are

$T = 2967$ K and $X = 0.984$

The pressure will then be obtained from

2-11 (continued, 2)

$$p(T) = p_1(T_2/T_1) = 10(2967/298) = 99.6 \text{ atm}$$

With a safety factor of 4, the tank should be able to withstand a pressure of essentially 400 atm.

For spherical containers, Table 12-10 provides a relation to obtain the minimum wall thickness t of the tank as

$$t = \frac{P r_i}{2 S E_J - 0.2P}$$

Where P is the maximum allowable pressure, r_i the inside radius, S the maximum allowable working stress, E_J the efficiency of the welded joint. Assume that the carbon steel used in the tank is SA-442 Grade 55 with an S value of 94.7 MPa and that the welds are butt joints with complete penetration and spot examined resulting in a weld efficiency of 85%.

The diameter of the spherical tank was not given. Assume inside diameters of 1, 2, and 5 meters. For a 1-m diameter sphere

$$t = \frac{(400)(1.013\times10^5)(0.5)}{2(94.7\times10^6)(0.85)-(0.2)(400)(1.013\times10^5}$$

$$= 0.1324 \text{ m} \quad \text{or} \quad 5.21 \text{ in}$$

For 2 and 5 m inside diameters, the minimum thicknesses are $\underset{\text{Answer}}{\underline{\underline{0.265 \text{ m and } 0.662 \text{ m}}}}$.

The results show the advantage of having a properly designed relief valve and burst disk assembly closely tied in with the reactor containing the decomposition reaction of acetylene. It would be imprudent to recommend a spherical container with a wall thickness greater than 0.0125 m because of the reactor costs involved.

2-12

The rate at which energy is produced in a reactor is given by the relation

$$\frac{dq}{d\theta} = \Delta H_{rx}\, k_o C V e^{-(E/RT)}$$

where $dq/d\theta$ is the heat liberated over time, ΔH_{rx} the heat of reaction, C the concentration of the reactant, V the volume of the reactor, E the activation energy, R the gas constant, and T the temperature.

When the reaction is exothermic, heat must be removed from the reactor to keep the temperature from increasing beyond control. The heat transfer rate can be written as

$$\frac{dq}{d\theta} = UA(T - T_o)$$

where U is the overall heat transfer coefficient, A the heat transfer area, and T_o the coolant temperature. If heat can be removed as rapidly as heat is generated by the reaction, the reaction can be kept under control. Thus, at steady state

$$\Delta H_{rx} k_o C V e^{-(E/RT)} = UA(T - T_o)$$

To simplify the relation, let $B = \Delta H_{rx} k_o CV$ and $K = UA$. Thus,

$$B e^{-(E/RT)} = K(T - T_o)$$

or $$B e^{-(E/RT)} - K(T - T_o) = 0$$

The maximum temperature where the steady-state energy balance is valid can be found by setting its derivative with respect to temperature equal to zero. Doing this, resuts in

$$B e^{-(E/RT)} (E/RT_c^2) - K = 0$$

where T_c is the critical or maximum temperature. However,

$$B e^{-(E/RT)} = K(T_c - T_o)$$

at the maximum temperature. Substituting this into the previous equation and solving for K gives

2-12 (continued)

$$K(T_c - T_0)(E/RT_c^2) = K$$

or $(T_c - T_0) = RT_c^2/E$ which can be used to find T_c

Substituting values

$$(T_c - 288.1) = 1.987\,T_c^2/28{,}000$$

$$T_c = 294.3 \text{ K}$$

This means that if the temperature in the reactor rises above 294.3 K (21.1°C), the reactor will begin to generate energy faster than it can be removed and the reaction will run away. Note that only a 6.2°C rise in temperature in the reactor can be tolerated.

Answer

For safe operation of the reactor some or all of the following actions can be taken:

1. Increase the size of the heat exchanger
2. Inject an inhibitor into the reactor or quench the reaction if it is ready to go out of control
3. Reduce the feed to the reactor
4. Provide adequate venting for the reactor for emergency use
5. Add diluents to the reactor to reduce the reaction rate. (These will probably have to be separated from the product after the reaction is completed)

2-13

A system for transferring liquid chlorine from a chlorine storage container with pressurized chlorine gas is shown below. A feature of this design is a relief system that utilizes a duplicate burst system to simplify replacement of the discs with a mechanical interlock between the two burst systems. Bottom outlets have been avoided so that joints are not exposed to liquid chlorine and the only leakage will be of a gaseos nature.

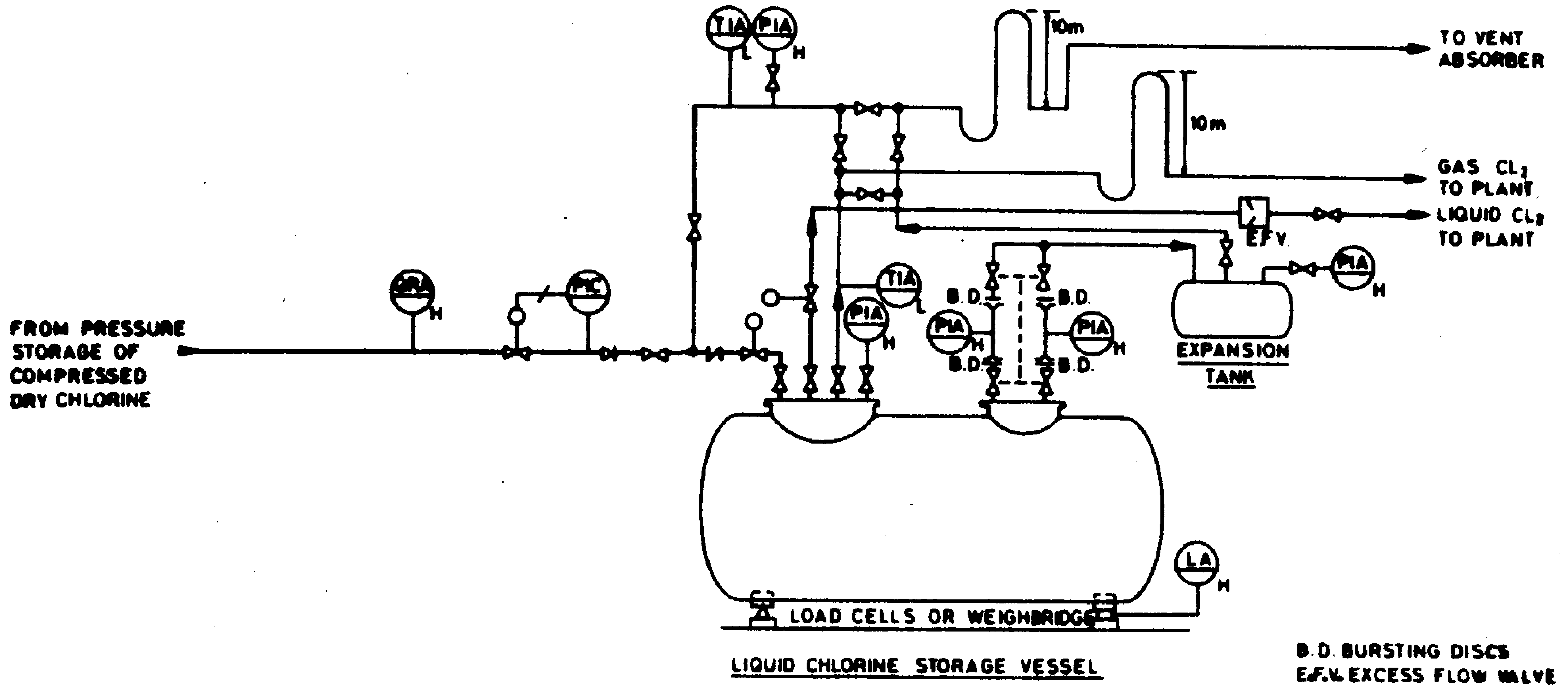
The severity of the leak is reduced by lowering the pressure within the storage container or system. A facility must be available to accomplish this pressure reduction during chlorine transfer as well. A waste Chlorine absorption unit that uses caustic or some simular chlorine consuming process is recommended. In addition, an expansion vessel is provided with an alarm to handle any pressure build-up in the container. The expansion tank is automatically vented to the absorption system (see figure).

The recompressed dry chlorine can be stored in a pressure storage receiver fitted with a relief valve set to discharge to the waste chlorine absorption unit at a predetermined pressure.

Pressurization with vaporized chlorine can function with a pressure storage tank that can utilize an external heater to vaporize the liquid chlorine. Control of the vaporization heater is tied to the pressure control system for the storage tank. Relief valves must permit over-pressurization to be relieved with excess chlorine vapor going to the waste chlorine absorption unit described above.

One possible instrumented arrangement for the chlorine vaporizer is shown in the second figure.

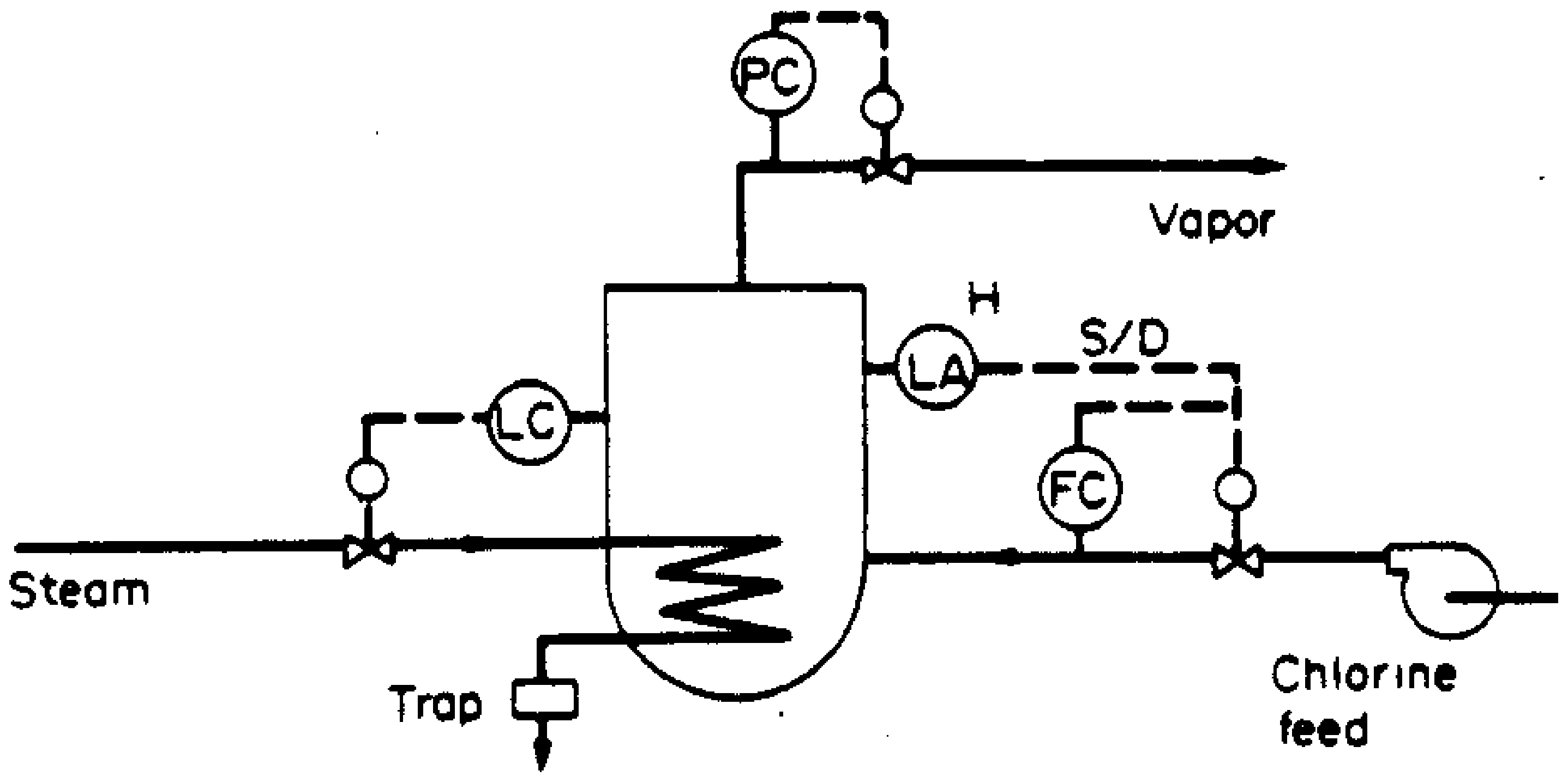
2-13 (continued, 1)



One possible arrangement with appropriate controls and alarm system for transferring liquid chlorine by using recompressed chlorine vapor.

Answer

2-13 (continued, 2)



Answer

2-14

Reducing the inventory of a hazardous material in a process includes, but is not limited to the following items:

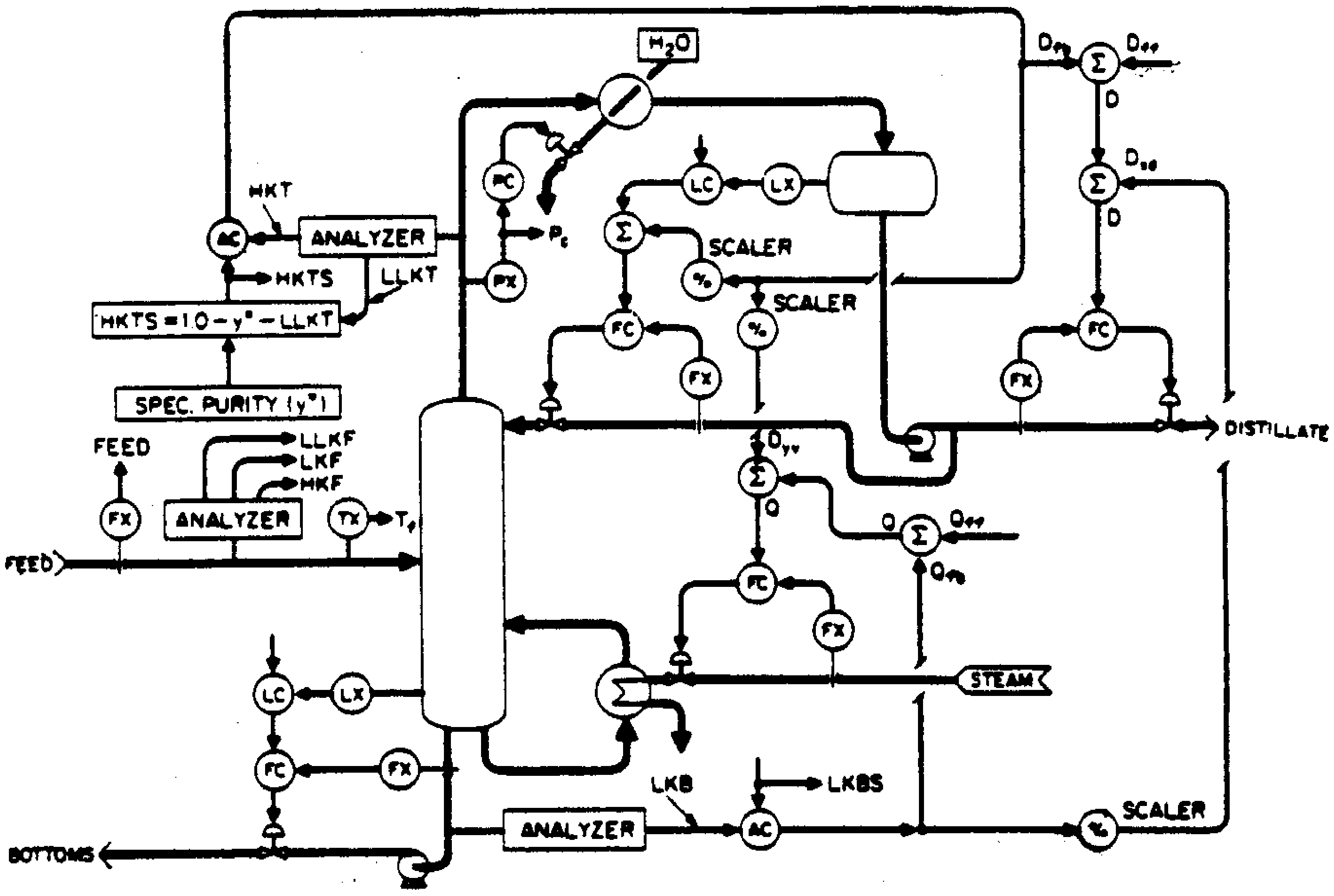
1. Investigate the possibility of producing the end product by another process that does not the use of a very hazardous material.
2. Determine whether the hazardous material may be reacted or combined before the distillation step to form a non-hazardous material. If the hazardous material is a reactant, recycle the stream back to the process to reduce the hazardous material content before the distillation step.
3. Depending upon how the hazardous material is used in the process, determine whether there are more efficient ways of absorbing, adsorbing, trapping, complexing, etc, the hazardous materials to avoid using a distillation operation that generally has considerably greater difficulty in attaining ppm concentrations in a product stream.
4. If elimination of the hazardous material is not possible, operate the distillation operation on a continuous basis. This will require less material being handled, smaller equipment used with possible elimination of the buffer storage, and improved chances of containment should there be a leak.
5. Improve the efficiencies of the column to reduce build-up of hazardous materials. Guidelines for such improvements are given by K.E. Nelson, Hydrocarbon Processing 69(3), 93-98 (1990).

<u>Answer</u>

A distillation column can be controlled in many ways. Generally, there is no one "best" way. Which one is to be prefered depends on the purpose of the separation, the types of controls available for each service, the variability of the feed, and the relative importance of the product recovery and product purity. Each case needs to be analyzed separately.

2-14 (continued)

A typical distillation column with appropriate instrumentation in which basic controls are assisted with predictive and analyzer loops is outlined in Perry's Chemical Engineers' Handbook, 6th ed., p. 22-118 and is reproduced below with the permission of McGraw-Hill.



Nomenclature Summary for Fig.

Symbol	Explanation
D_{fb}	Contribution of feedback control to distillate-flow set point
D_{ff}	Contribution of predictive control
D_{nd}	Contribution of noninteractive decoupler to distillate-flow set point
D_{yr}	Contribution of noninteractive decoupler to reboiler-heat set point
HKF	Heavy key feed-stream component
HKT	Heavy key overhead-product-stream-component concentration, weight percent
HKTS	Specified concentration of heavy key component in overhead product, weight percent
LKBS, x^*	Specified concentration of light key component in bottoms product, weight percent
LKF, z	Light-key-feed-stream-component concentration, weight percent
LLKF	Lighter-than-light-key-feed-stream-component concentration, weight percent
LLKT	Lighter-than-light-key-overhead-product-stream-component concentration, weight percent
Q	Reboiler heat, BTU/unit time
Q_{fb}	Contribution of feedback control to steam-flow set point
Q_{ff}	Contribution of predictive control to steam-flow set point
Q_r/F	Weight ratio, reboiler-heat-to-feed
x^*, *LKBS*	Specified concentration of light key component in bottoms product, weight percent
z, *LKF*	Light-key-feed-stream-component concentration, weight percent

2-15

HAZOP for Line L-102

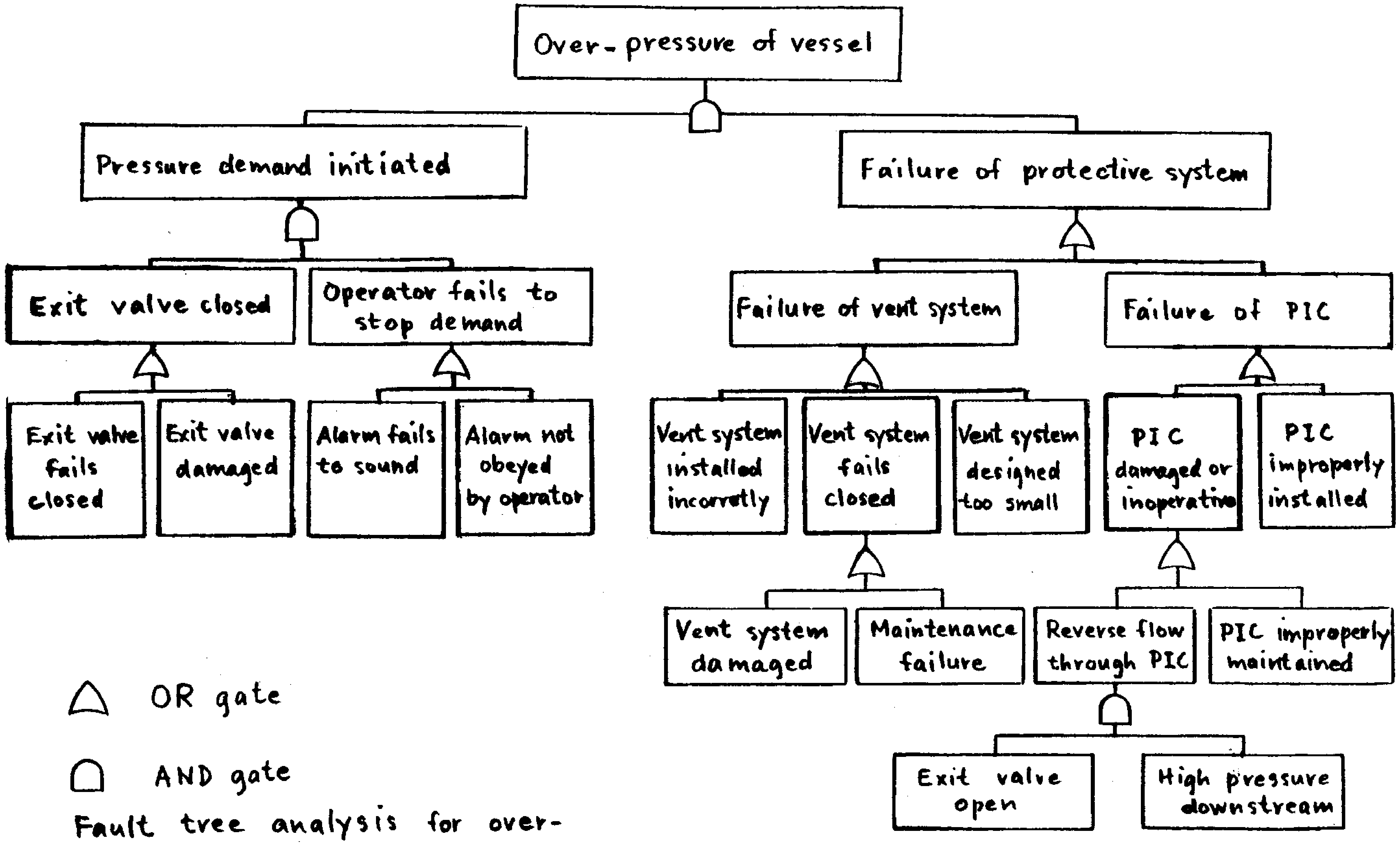
Guide word	Property	Possible cause	Possible consequence	Action required
More	Temperature	A	Pump seal failure, vapor lock	Install a feed-back line
Less	Temperature	Low ambient temperature	B	steam tracing
More	Flow	Line fracture	Spillage (possible large explosion)	C
		Pump seal failure	Spillage (possible small explosion)	D
		Control fault		Consider bypass
No	Flow	E	Shutdown	Low level alarm
		F	Shutdown	Automatic start-up of standby pump
Reverse	Flow	Pump failure	Backpressure on storage vessel	G
As well as	Impurities	H	Possible small detonation	Priming line

The missing items below complete the HAZOP study above:

A. Steam tracing operating incorrectly or downstream valve closed
B. Thermal stress on line
C. Install flow control with automatic shutoff
D. Flow diverted through standby pump capable of automatic startup
E. Tank empty
F. Pump failure
G. Relieve backpressure, divert flow through standby pump with automatic startup
H. Pump drawing in air or operator error

Answer

2-16



Fault tree analysis for over-pressurizing vessel

2-17

The important factors to be considered for plant location in the fertilizer industry are (1) raw materials, (2) markets, (3) transportation facilities, (4) power and fuel availability, (5) waste disposal, and (6) water supply. The order of importance varies with the fertilizer considered and to some extent to the size of plant contemplated. Analysis of this situation for ammonia, urea, and phosphate fertilizer requires a knowledge of the raw materials required and the most commonly used process for each fertilizer. A concise review of these requirements is given by G.T. Austin in Shreve's Chemical Process Industries, 5th ed. About 90% of US ammonia production uses natural gas as the feedstock. The reactions are

$$CH_{4(g)} + H_2O \xrightarrow[2860\ \text{kPa},\ 515°C]{\text{Ni catalyst}} CO_{(g)} + 3\,H_{2(g)}$$

$$CH_{4(g)} + \text{air} \xrightarrow[670°C]{\text{Ni catalyst}} CO_{(g)} + 2H_{2(g)} + N_{2(g)}$$

$$1/2\,N_{2(g)} + 3/2\,H_{2(g)} \xrightarrow[\text{pressure}]{\text{Fe catalyst}} NH_{3(g)}$$

The process requires large amounts of power and cooling water. Since natural gas is easily transported by pipeline, the plant need not be located near the gas well. Rather, it is more profitable to locate the plant close to the markets. Thus, the latter is the most important factor in plant location. Adequate power and water supply would also rank quite high. As the size of the plant increased, more consideration would have to be given to transportation facilities since the markets would be from the plant. Adequate supply of natural gas would also become important with plant size.

The commercial processes in current use for the manufacture of urea are based on two reactions, namely

$$CO_{2(g)} + 2NH_{3(l)} \xrightarrow[185°C - 193°C]{22\ \text{MPa}} NH_4CO_2NH_{2(l)}$$

2-17 (continued)

$$NH_4CO_2NH_2 \rightleftharpoons NH_2CONH_{2(l)} + H_2O_{(l)}$$

This process requires approximately 1/4 the power and cooling water of that required for the ammonia process per unit weight of fertilizer produced. Since ammonia is one of the principal raw materials required for the process, it is much more important to consider the source of the raw materials in a selection of the plant site for the production of urea. This factor will be the most important consideration even with an increase in plant size. However, since the fertilizer markets are not as large as for the raw material, careful consideration must be given to markets and the transportation necessar to reach those markets.

Phosphate fertilizer is obtained from phosphate rock ore. The accidulation of phosphate rock to produce superphosphate fertilizer can probably best be described by the following reaction:

$$CaF_2 \cdot 3Ca_3(PO_4)_2 + 7\ H_2SO_4 + 3H_2O \longrightarrow 3CaH_4(PO_4)_2 \cdot H_2O + 2HF + 7CaSO_4$$

Since the phosphate rock must be concentrated after it is mined, plant location is determined by the location of the phosphate rock ore regardless of plant size. The aspects of waste disposal of the non-phosphate rock is directly related to the mining operation. Markets directly affect plant size.

<u>Answer</u>

2-18

The diluted waste sample contains 15g of waste and 485g of dilution water. The total oxygen available for consumption by the waste is

$$\text{g of } O_2 = (485)(9)/10^6 = 4.365 \times 10^{-3}\,g$$

The final oxygen content of the diluted sample is

$$\text{g of } O_2 = (500)(4)/10^6 = 2.0 \times 10^{-3}\,g$$

The oxygen depletion due to the 15 g waste sample is

$$O_2 \text{ depletion} = 4.365 \times 10^{-3} - 2.0 \times 10^{-3} = 2.365 \times 10^{-3}\,g$$

The BOD_1 defined as the oxygen depletion in parts oxygen consumed per 10^6 parts of waste is then

$$BOD = (2.365 \times 10^{-3})(10^6/15) = \underline{\underline{158 \text{ parts } O_2/10^6 \text{ parts waste}}}$$

Answer

2-19

The solution to this problem involves some of the same safety aspects as considered in the solution of Problem 2-13.

The most common method to vaporize chlorine is to use steam in a heating coil immersed in the chlorine liquid in the vaporizer. A schematic with typical instrumentation is shown in the solution of Problem 2-13 and will not be repeated here.

The disadvantage of using hot water to vaporize the chlorine is the need for a larger coil and a greater potential for freezing of the coil (NBP of Cl_2 is −34°C). Economics is probably the major deterrent for not using a closed circuit heating with a heat transfer fluid that is inert to chlorine.

The use of a heating coil immersed below the liquid level in a chlorine vaporizer is the preffered arrangement to accomplish the vaporization. Not only is the coiled tube less expensive, it is also less susceptible to leaks and failures.

Answer

2-20

The net heat transfer flux to a storage tank during a fire comes from radiation and convection attributed to the fire that surrounds tank and can be approximated by

$$q_{net} = q_r + q_c + q_{rr}$$

where q_r is the radiation flux from the fire, q_c the convection flux from the fire, and q_{rr} the radiation flux back from the tank. The radiation fluxes are proportional to the differences in the fourth powers of the absolute temperatures for the radiating and absorbing surfaces. The convection flux can be estimated of natural convection heat transfer. Since the surface of interest in this problem is large and vertical, the relations for natural convection on vertical plane surfaces are the most appropriate ones to use. The flame can be assumed to have transport properties equal to that of hot air in the flame and the fact that most of the gas in the flame is nitrogen. Flame temperatures vary from fuel to fuel, but a temperature of about 1500K will provide a reasonable average value for most situations.

If the benzene tank, assumed to be a vertical cylinder, is to store 1600 m^3, its minimum height is given by

$$V = \pi D_i^2 Z/4$$

$$Z = 4V/\pi D_i^2 = (4)(1600)/\pi(15)^2 = 9.05 \text{ m}$$

Storage tanks generally have an ullage volume of 10% or more to take care of liquid expansion and provide additional volume for some vaporization to occur without appreciably increasing the tank pressure. Heat transfer to the surface area surrounding the ullage space will be minimal compared to the surface area wetted by the full tank contents. Thus, only the latter area will be used in the heat transfer calculation. Also assume that the temperature of the wetted wall to be at the boiling point temperature of the fluid being stored in the tank. There will be a temperature gradient through the tank wall, but this will be small and have minimal effect on the re-

2-20 (continued, 1)

quired venting rate during a fire.

$$q_r = 95{,}000 \text{ W/m}^2$$

is given in the problem statement and assumes that the external fire completely surrounds the storage tank. The convective flux from the fire is given by

$$q_c = h(T_f - T_w)$$

where T_f is the flame temperature and T_w is the wall temperature, assumed to be the same as the normal boiling temperature of the fluid in the tank. The heat transfer coefficient in US customary units for natural convection along a vertical surface may be obtained numerically from

$$Nu = \left\{ 0.825 + \frac{0.387\, R_a^{1/6}}{\left[1 + \left(\frac{0.492}{P_r}\right)^{9/16}\right]^{8/27}} \right\}^2$$

as given in <u>Perry's Chemical Engineers' Handbook</u>, 6th ed., where $Nu = hL/k$, $Pr = C_p\mu/k$, $Ra = Gr\,Pr$, and $Gr = L^3 \rho^2 g \beta \Delta T/\mu^2$. Since this calculation is rather lengthy and natural convective heat transfer coefficients for air generally have values of 5 to 10 W/m²·K, we will assume an average value of 7.5 W/m²·K and only calculate a better value if analysis warrants it.

Using this assumption

$$q_c = (7.5)(1500 - 353) = 8600 \text{ W/m}^2\cdot\text{K}$$

The heat flux radiated back from the tank surface to space can be estimated from

$$q_{rr} = \epsilon \sigma T_w^4 \qquad \text{assume } \epsilon = 1.0$$

$$= (1.0)(0.972 \times 10^{-8})(353)^4 = 150 \text{ W/m}^2\cdot\text{K}$$

From these calculations

$$q_{net} = 950{,}000 + 8600 - 150 = 103{,}450 \text{ W/m}^2\cdot\text{K}$$

2-20 (continued, 2)

The vapor boil-off rate is then

$$\dot{m} = (q_{net}) A / \Delta H_v \qquad \Delta H_v = \text{heat of vaporization}$$

$$= (q_{(net)}) \pi D Z / \Delta H_v$$

$$= (103,450) \pi (15)(9.05) / 3.943 \times 10^5 = 111.9 \text{ kg/s}$$

At the NBP of benzene, the vapor density is 2.755 kg/m^3. Thus

$$\dot{m}_v = \dot{m} / \rho = 111.9 / 2.755 = 40.65 \text{ m}^3/\text{s}$$

The vent area needed for this volumetric flow rate can be determined by use of an orifice equation of the form

$$A = \frac{\dot{m}}{0.61 [(2) \Delta p \rho]^{1/2}} \qquad \Delta p = 250 \text{ Pa}$$

$$= \frac{111.9}{(0.61)[(2)(250)(2.755)]^{1/2}} = 4.94 \text{ m}^2$$

The area for the vent is much too large for the tank. Therefore a weak seam roof should be specified

Answer

Several assumptions were made in the solution of this problem and need to be verified. First, the convective flux is only 9% of the total heat flux. Even if it were changed ± 100%, the overall result would only be changed slightly. Thus, an exact calculation of the heat transfer coefficient would not change the final conclusion.

There is some resistance to heat transfer through the walls of the tank. For a tank wall of 0.0125 m, the temperature gradient is

$$\Delta T = q_{(net)} t_w / k_w = (103,450)(0.0125)/45 = 28.7°\text{C}$$

This will have a negligible effect on the results, and thus this assumption is acceptable. However, there is also a thermal resistance with the film on the inside of the tank. Typical heat transfer coefficients are on the order of several thousand $W/m^2 \cdot K$. Assuming a value of

2-20 (continued, 3)

2,000 $W/m^2 \cdot K$, the ΔT across the film is

$$\Delta T = q_{(net)}/h = 103{,}450/2000 = 51.7\,°C$$

The combination of these two resistances only reduces the convective flux rate by approximately 6%. This change would have negligible effect on the final answer.

As a footnote, the NFPA Fire Protection Handbook and API 650 all recommend the following equation for the net heat flux to a tank subjected to an external fire:

$$q_{(net)} = 43{,}200\, A^{0.82}$$

where $q_{(net)}$ is in watts and A is the wetted area in m^2. For this problem

$$q_{(net)} = 43{,}200\,[\pi D_t Z]^{0.82} = 43{,}200\,[\pi(15)(9.05)]^{0.82}$$
$$= 6.19 \times 10^6\ W$$

from this

$$\dot{m}_V = q_{(net)}/\Delta H_v \rho$$
$$= 6.19\times10^6/(3.943\times10^5)(2.755) = 5.6\ m^3/s$$

The major difference in results is that the heat flux in the problem statement applies to a fire that completely surrounds the tank. Generally fires are seldom large enough to surround large tanks. In addition a soot layer builds up on the cool tank walls which further restricts the heat transfer. Thus, the total heat flux is reduced substantially. However, the venting rate even under these conditions will be large and a tank with a weak roof seam would be recommended.

3-1

The production of formaldehyde still relies on the chemical process developed nearly 50 years ago. Nearly 70% of new installations in the US utilize a metal oxide catalyst to form formaldehyde by the oxidation of methanol

$$CH_3OH + \frac{1}{2} O_2 \rightarrow HCHO$$

Process 1

In this process shown above, vaporized methanol is mixed with air and recycle tail gas and passed through catalyst filled tubes in a heat exchanger reactor. Heat released by the exothermic reaction is removed by vaporization of a high-boiling heat transfer fluid on the outside of the tubes. Steam is normally produced by condensing the heat transfer fluid. Product leaving the bottom of the reactor is cooled and then transferred to the base of an absorber. Formaldehyde concentration in the product is adjusted by controlling the amount of water added to the top of the absorber. A product of 1 to 55% formaldehyde can be produced. Byproducts in the reaction include CO, CO_2, dimethylether, and formic acid. The latter is removed by ion exchange. Plant yields between 88 – 92% are reported.

Reference: "Kirk-Othmer Encyclopedia of Chemical Technology", Vol. II, p. 929, J. Wiley, New York, 1994.

Process 2

A feed mixture is generated by sparging air into a pool of heated methanol and combining the vapor with steam. The mixture passes through a superheater to a catalyst bed of silver crystals or layers of silver gauze. The reactor product is rapidly cooled in a steam generator followed by a water-cooled heat exchange and then fed to the bottom of an absorption tower. Most of the methanol, water, and formaldehyde is condensed in the water-cooled bottom section of the absorber. Almost complete removal of the remaining methanol and formaldehyde from the tail gas occurs at the top of the tower by countercurrent contact with the process water. Absorber

3-1 (continued)

bottoms go to a distillation column where methanol is recovered for recycle to the reactor. The bottoms stream from the distillation tower is usually sent to an anion exchanger which reduces the formic acid to specifications. The product contains up to 55% formaldehyde and less than 1.5% methanol.

Reference: J.H. Martin and M.T. Butter, *Oil & Gas J.* **72**, 71(Mar. 11, 1974).

Process 3

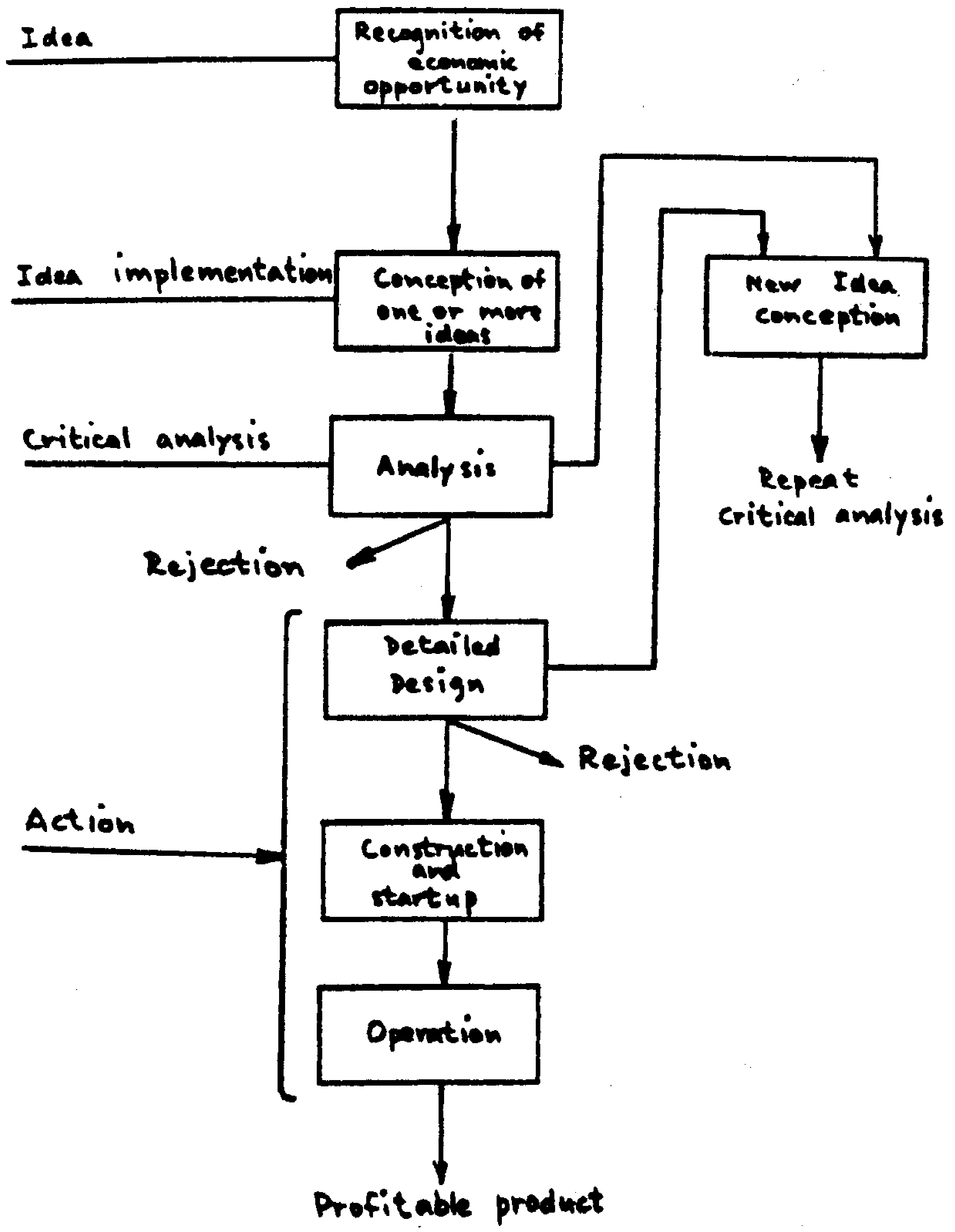
Another process to produce formaldehyde involves the partial oxidation of methane. The incentive for such a process is the reduction of raw material costs by avoiding the capital and operating expense of initially producing the methanol from methane. In this process methane gas is mixed with excess air, preheated to around 600°C and fed into a vertical, tubular reactor with a conical bottom, filled with ceramic packing saturated with a $K_2B_4O_2$ solution. The conversion of methane depends on the temperature of the feed mixture and its content of nitrogen oxides. Exit gases are rapidly cooled to 400°C in a tubular shell condenser with water. The gas is further cooled to approximately 33°C in a heat exchanger and transferred to a scrubber where formaldehyde is absorbed in a dilute solution of formaldehyde solution to provide formaldehyde of the desired concentration. Since energy requirements are quite high, the processing plant is best coupled with a cheap source of power.

Reference: M. Brown and N. Parkyns, *Catal. Today* 5(3), 305 (1991).

J. Hargreaves, G.Hutchings, and R. Joyner, *Nature* **348**, 28 (Nov. 1990).

3-2

A general outline showing the sequence of steps in the development of a formaldehyde plant is shown below:



3-2 (continued)

Recognition of economic opportunity – this can come from a variety of sources. Technical services may recognize a customer's need in a resin or adhesive. Market research may uncover another use of formaldehyde as a feed product or intermediary requirement in a process.

Conception of one of more ideas – again this can come from a variety of sources, namely, the chemist, process engineer, plant operator, technical service personnel, maintenance crew, etc. The idea can be one of an entirely new process like the utilization of natural gas for a raw material (see Problem 3-1) rather than methanol. On the other hand, the idea can involve a change in operating conditions and arrangement of equipment while using the same raw materials as demonstrated by the Formox process developed by Reichhold Chemicals (see *Hydrocarbon Process.*, **44**(11), 215 (1965) or the Montecatini process (see *Hydrocarbon Process.*, **44**(11), 216 (1965).

Analysis – this step requires the greatest technical interaction on the part of the design engineer. For the production of formaldehyde this step includes the following considerations:

1. Raw materials (availability, quantity, quality, cost)
2. Thermodynamics and kinetics of the chemical reaction of, for example, the Formox process

   $$2CH_3OH + O_2 \rightarrow 2HCHO + 2H_2O$$

   in terms of equilibrium, yields, rates, optimum conditions, etc.
3. Facilities and equipment available and what must be purchased
4. Estimation of production costs and total investment for various alternative designs
5. Profits (probable and optimum per kilogram of formaldehyde as well as annually, return on investment)
6. Materials of construction
7. Safety considerations

3-2 (continued)

8. Markets (present and future supply and demand, present uses, new uses, etc.)
9. Competition (overall production statistics, comparison of various manufacturing processes, product specification, e.g. 37% formaldehyde)
10. Properties of formaldehyde, properties of byproducts, disposal of waste products – this can be rather difficult in the case of the formaldehyde process.
11. Sales and sales service
12. Shipping restrictions, labeling, container selection
13. Plant location
14. Patent situation and legal restrictions

The analysis step may indicate that there is insufficient data which may require additional research. The analysis step involves the development of a preliminary design. This may lead to rejection of the entire project or it may initiate the development of additional new ideas. The latter should undergo further critical analysis to determine whether these, in turn, are feasible and should be pursued further or also rejected.

Detailed design – the following factors should be established within narrow limits before a detailed design is performed:

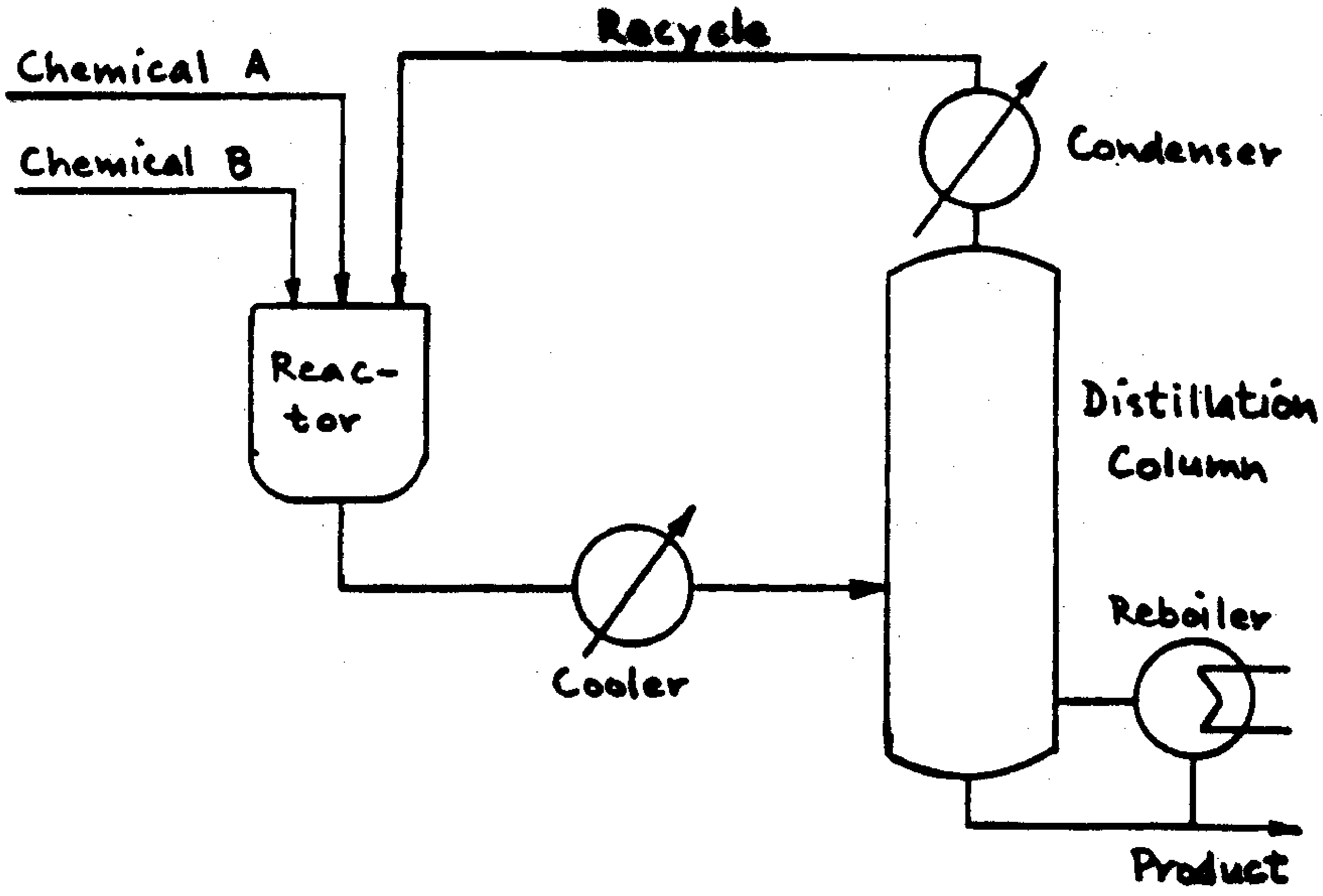
1. Selected process
2. Material and energy balances
3. Temperature and pressure ranges
4. Raw material and product specifications
5. Yields, reaction rates
6. Materials of construction
7. Utilities requirements
8. Plant site

If the detailed design indicates a profitable venture that meets the return criteria set forth by company managers, a recommendation will be made to go ahead with the construction if funds are available.

3-2 (continued)

Construction – the final design should be developed in such a manner that items requiring the longest time for completion can be started first. Careful coordination between the design engineer and the contractor will help minimize construction delays and assist with more orderly plant startup. Detailed operating procedures need to be developed not only for startup procedures, but also for eventual continuous operation. During operation, engineers and operating personnel should be encouraged to continuously monitor the process to see where the process can be improved or simplified.

3-3



Information needed for the design:

Raw materials – chemicals involved, stream capacities, thermal properties of materials, physical properties of materials, temperature, pressure, material concentrations, environmental hazards

Product – some considerations listed for the raw materials above, product purity, build-up of byproducts

Reactor – rate expression, conversion under various operating conditions, heats of reaction, thermal and physical properties of entering and exiting materials, temperature, pressure, stream capacities

Cooler – quenching temperature, vapor-liquid equilibria, heat load, thermal and physical properties

3-3 (continued)

Distillation column – vapor-liquid equilibria, heat of vaporization, thermal and physical properties, stream capacities, temperature, pressure, reflux ratio, plate efficiency, stream capacities

Condenser and reboiler – heat of condensation and vaporization, heat loads, thermal and physical properties, temperature, pressure, heat transfer coefficient, stream capacities

3-4

Acetylene is currently produced by the thermal cracking of hydrocarbon feeds. This thermal cracking requires considerable quantities of energy since the reactions are strongly endothermic. Further dissociation into carbon and hydrogen of all hydrocarbons begins at relatively low temperatures. Thus, the reactions must be allowed to proceed very rapidly and quenched very quickly to stable temperatures.

One method of distinguishing between processes for acetylene production is by the manner in which the energy for cracking is supplied.

1. Regenerative method

This method involves periodic cycles of heating a hot surface of ceramic materials and then cracking the hydrocarbons on these hot surfaces. An example of this method is the Wulff process. High purity acetylene is possible from a feedstock of ethane to gas oil. Byproducts are an aromatic distillate fraction and a hydrogen-rich off gas. The reaction is a vapor phase reaction in the presence of steam. Low conversion between 12 to 32 wt% is a major drawback of this method. Also, startup and shutdown are quite long compared to some of the other processes.

2. Partial oxidation

Acetylene is obtained from the incomplete combustion of methane. Oxygen and natural gas are reacted and about one-third of the entering methane is cracked and the rest is combusted to provide the necessary heat for the cracking process. The yield based on the carbon in the natural gas is usually about 31 wt%. The process generates considerable soot (carbon) that must be removed continuously for good operation. The process is licensed by BASF.

3. Two-step procedure

This process involves the generation of combustion gases at about 1500K and subsequent injection of hydrocarbons into these hot combustion gases. The feedstock normally is naptha cuts

3-4 (continued)

to middle distillates. High purity acetylene is obtained with a high yield of 50 to 54 wt% from this feedstock. The process generally is identified as the Hoechst HTP process.

4. Submerged-flame process

In this process the reaction between hydrocarbons and oxygen occurs in a submerged-flame burner where the burners remain submerged in oil during operation. The burner projects a flame into which oil and oxygen are fed. Combustion and cracking of the oil takes place at the boundaries of the flame. The gases are rapidly quenched by the surrounding oil which is continuously cooled. The feed is crude oil rather than natural gas. The process is designed to recycle the soot and combust it in the burner. Product yields are in the same range as those obtained for processes that use the light hydrocarbons for feed stock.

Since the flowsheets for all of these four processes are readily available in the literature along with useful information on the operating details and procedures, they will not be reproduced here. Some of the earlier references to the processes listed above provide much more useful information than is available from later references. Accordingly, these references are listed rather than the more recent references.

1. Regenerative method: *Hydrocarbon Process.* **46**(11), 139 (1967)
2. Partial oxidation: *Hydrocarbon Process.* **48**(11), 142 (1969)
3. Two-step procedure: *Hydrocarbon Process.* **46**(11), 138 (1969)
4. Submerged-flame process: *Chem. Eng.* **76**(3), 82 (1969)

The choice of one process over another is largely dependent upon the projected needs. The availability of feed stocks, the desired product purity, and byproduct marketability will be overriding considerations. As noted above, one basic difference is the form of energy input. The availability of one form of energy over another could make one process more attractive than another. The licensing and fees requirements could also provide advantages of one process over another. The principal design problems involve the chemical reactions and energy requirements.

3-4 (continued)

The latter will be a major part of the operating costs and will be reflected in the sales price. Material and energy balances around the reactor or burner will be necessary. However, complete kinetic data for these complex reactions will probably not be available. In the case of these well-known processes, the information needed would be available from the licensor.

The other area of concern in all of these processes is in the separation reaction. Most require the use of selective solvents. The mass transfer calculations for these solvents would be needed. Again, data would be available from the licensor, though approximations can be made with today's sophisticated computer software.

There are many similarities in the various separation techniques. Thus, comparison of the processes can provide economic information in some cases without making an actual detailed cost estimate.

3-5

Some of the most important and widely employed processes for ethylene production utilize thermal cracking of a hydrocarbon feed stock in the presence of a steam atmosphere. As noted in the preliminary design in Chapter 3, the ethylene process based on thermal cracking consists basically of four distinct process units, namely (1) thermal cracking and quenching, (2) compression and acid gas removal, (3) subcooling and product separation, and (4) refrigeration. Selection of a feed stock for an ethylene plant is primarily dependent on feed stock availability, relative feed price, yield structure, plant construction cost, as well as the sale price for the byproducts. The thermal cracking of ethane and propane by the fixed tubular heater is the most common process in the United States, and the thermal cracking of naptha by the fixed tubular heater is the most common process in Europe and Japan.

The capital cost of an ethylene plant increases as the molecular weight of the feed stock increases since the ethylene yield decreases with increasing molecular weight. In theory, an ethylene plant can be designed to process a full range of feed stocks from ethane to gas oil. However, in practice, complete feed stock flexibility to cover so broad a range is neither technically practical nor economically attractive because of various constraints in equipment design.

Distribution of product yields depends on the type of feed stock, hydrocarbon partial pressure, residence time, as well as cracking severity or coil outlet temperature. Typical yield distribution patterns for a variety of feedstocks is shown in the table below. The cracked gas recovery section in an ethylene plant is complex and allows a variety of possible flow schemes. In the separation section it is necessary to separate hydrogen and methane fractions, high-purity ethylene and propane, ethane and propane fractions, C_4s and gasoline. The design must provide for removal of acetylene and methyl acetylene/propadiene (MAPD). For the separation process, there are maximum temperature limits to prevent polymerization fouling, and minimum temperature limits to prevent hydrocarbon freezing or hydrate formation. Within the separation process, there are several possible variations. Acetylene can be removed from the cracked gas stream in the

Typical pyrolysis yields of ethylene plant feedstocks (in mass percent).

Feedstock	Ethane	Ethane	Propane	Propane	n-Butane	Isobutane	n-Pentane	Isopentane	Naphtha	Naphtha	Kerosene	Gas Oil	Gas Oil (Vac)
Conversion%	50	60	75	90	95	92	95	95					
Severity									Medium	High	High	High	High
Comp. Yield													
H_2	3.06	3.55	1.08	1.29	1.0	1.39	0.8	0.9	0.63	0.8	0.63	0.65	0.47
CH_4	2.6	4.2	18.83	24.7	21.8	23.6	17.52	20.6	11.7	15.3	11.4	10.6	7.9
C_2H_2	0.12	0.25	0.17	0.33	0.4	0.46	0.55	0.42	0.3	0.75	0.37	0.35	0.2
C_2H_4	41.65	48.2	25.0	34.5	35.8	10.3	36.5	17.95	24.0	29.8	23.2	24.0	19.5
C_2H_6	50.0	40.0	5.15	4.4	5.1	0.73	5.6	3.2	3.9	3.75	3.4	3.25	2.83
C_3H_4	0.1	0.02	0.22	0.34	0.55	2.46	0.95	1.7	0.76	1.1	0.85	1.0	0.38
C_3H_6	0.89	1.11	18.86	13.96	16.4	21.2	19.8	19.0	15.75	14.1	13.8	14.45	11.3
C_3H_8	0.14	0.17	25.0	10.0	0.15	0.33	0.55	0.55	0.41	0.25	0.45	0.45	0.36
C_4H_6	0.5	1.07	1.05	2.65	3.4	1.94	3.7	4.6	4.6	4.85	4.4	4.7	5.5
C_4H_8	0.25	0.21	1.28	1.0	1.7	16.7	2.6	14.38	6.85	4.2	4.0	4.5	4.88
C_4H_{10}	0.35	0.27	0.07	0.05	5.0	8.0	0.1	0.1	1.0	0.25	0.1	0.1	0.09
C_5s	0.2	0.27	1.43	1.81	1.65	2.12	6.1	7.5	5.0	2.3	3.0	3.3	5.0
C_6H_6	0.2	0.48	1.09	2.2	2.58	3.42	1.8	4.1	3.15	6.3	7.2	5.8	4.16
C_7H_8	0.03	0.06	0.33	0.48	1.12	1.51	0.6	1.25	4.4	4.9	4.0	3.15	2.5
C_8H_8					0.45	0.67	0.1	0.25	2.65	2.0	2.2	1.2	1.28
C_8H_{10}							0.06	0.15	0.52	0.7	0.75	0.7	0.5
C_6-C_8 PON*		0.14	0.14	1.44	1.2	3.18	0.1	0.1	8.83	2.0	1.9	1.8	2.0
C_9-200^0 C							1.48	1.75	3.75	2.8	4.85	2.4	2.0
Fuel oil			0.3	0.85	1.7	1.99	1.09	1.5	1.8	3.8	13.5	17.6	29.15

* PON (Paraffin, Olefin, Naphthene).

3-5 (continued)

presence of the hydrogen produced in the furnaces or from a concentrated demethanized and depropanized C_2 cut. Such process arrangements are identified as the front-end demethanizer, the front-end deethanizer, and the front-end deproponizer scheme. They are briefly described below.

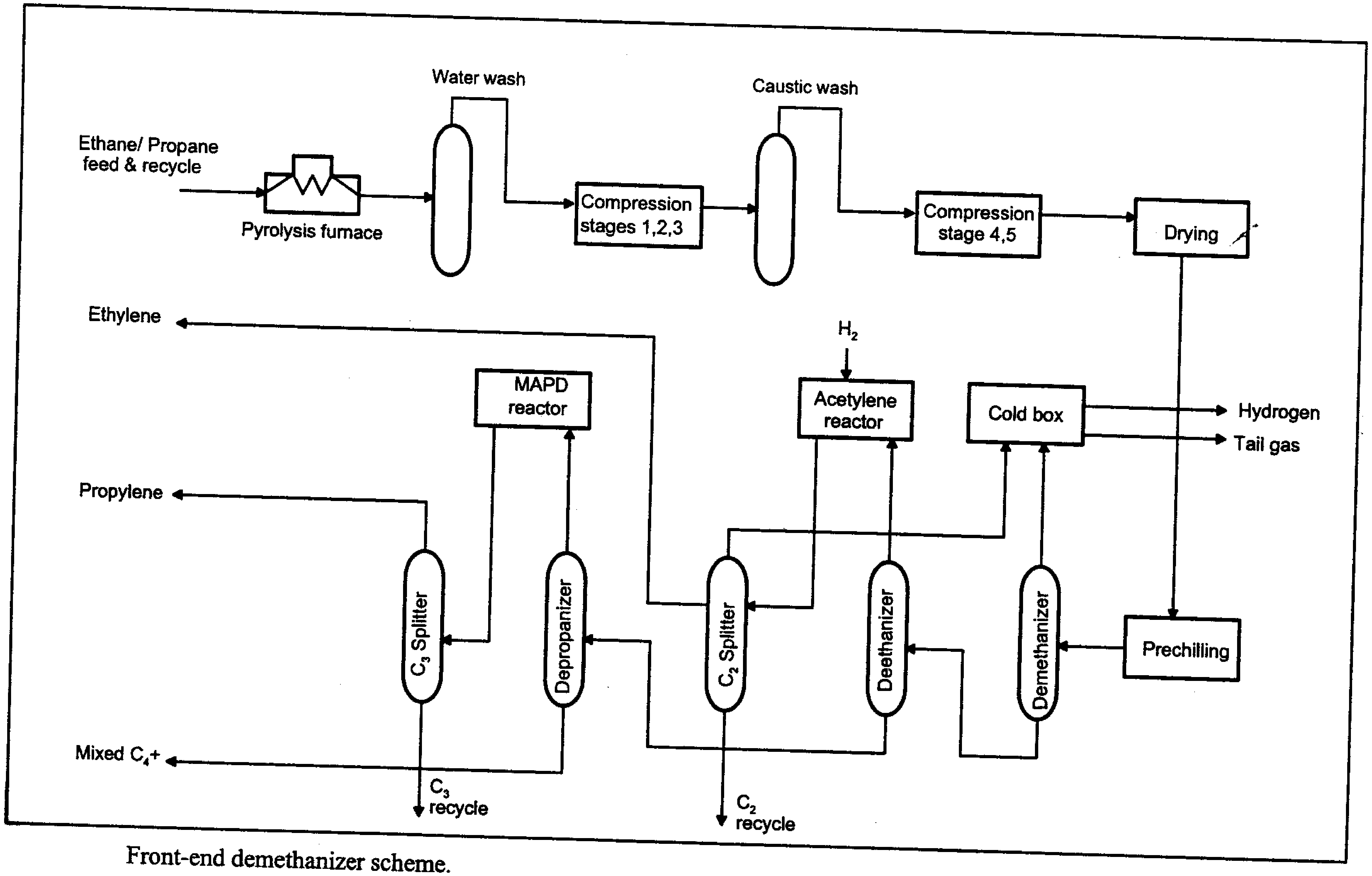
Front-end demethanizer

In this scheme the dried cracked gas is chilled and sent to the demethanizer/cold-box system as shown in the attached figure. The demethanizer bottoms are sent to the deethanizer. The overhead from the deethanizer is hydrogenated to remove the acetylene. From the acetylene converter, the ethylene-ethane stream passes to the C_2 splitter. The deethanizer bottoms are sent to a depropanizer which separates the C_3 components from the heavier C_4 components. The overhead from the depropanizer is fed to a methyl acetylene/propadiene hydrogenation reactor. A C_3 splitter separator separates the propylene from propane. The bottoms stream from the depropanizer passes to a debutanizer where the C_4 fraction is separated as an overhead from the C_5 and higher components.

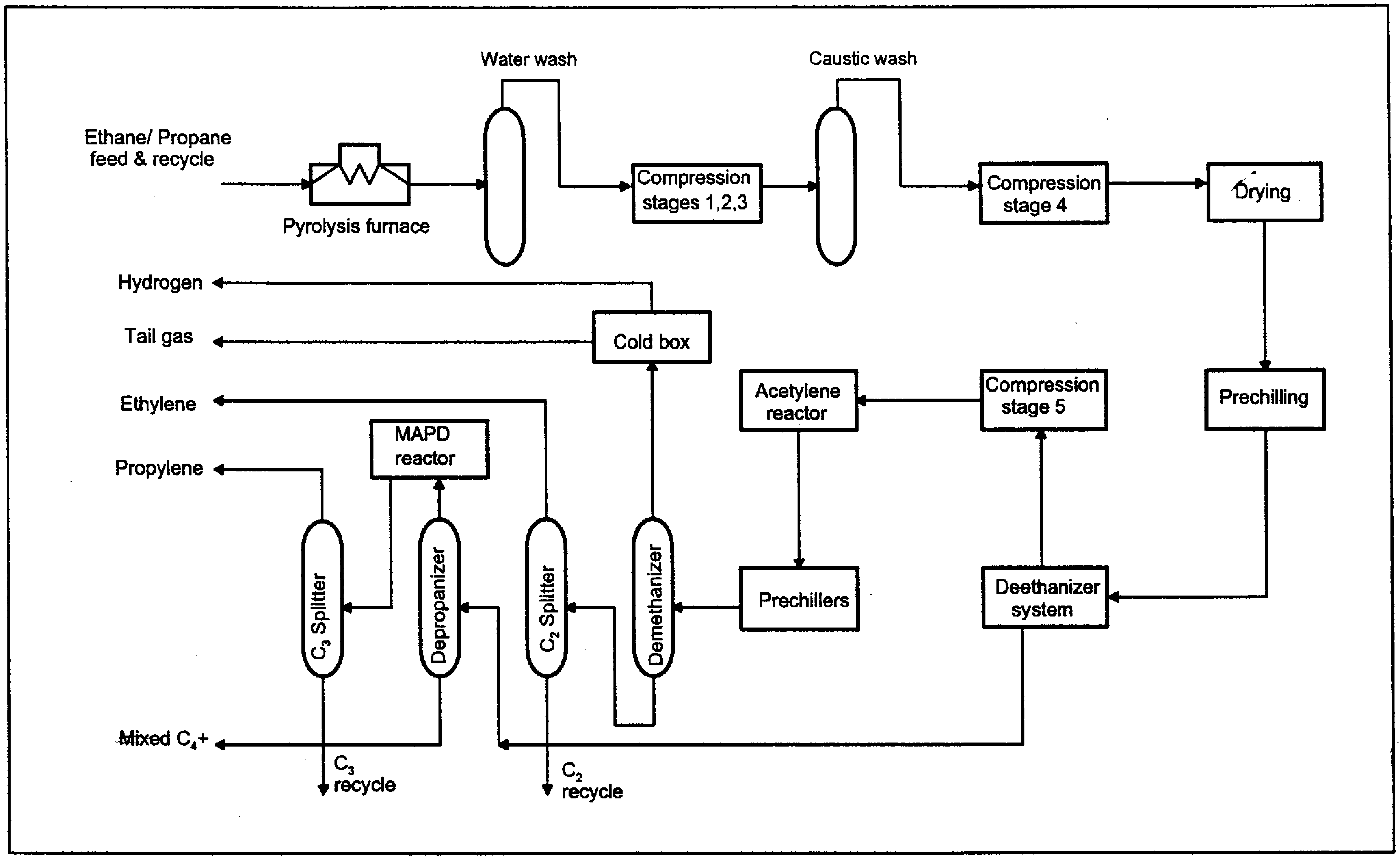
Front-end deethanizer

As noted in the figure showing the front-end deethanizer, the cracked gas is dried and chilled before it is sent to the deethanization section. Overhead from the deethanization section is compressed and sent to a front-end acetylene reactor where acetylene is hydrogenated to ethylene and ethane. Following the acetylene reactor, the C_2 and lighter components are chilled and sent to the demethanizer which separates methane and hydrogen from the heavier components. Hydrogen and methane are separated in the cold-box, while the demethanizer bottoms are fractionated in the C_2 splitter to produce ethylene product and ethane for recycle. The bottoms from the deethanizer are depropanized, and the mixed C_3s are sent to the methyl acetylene/propadiene reactor for hydrogenation. The C_3 reactor effluent goes to a C_3 splitter to produce polymer grade propylene and propane for recycle.

3:15



Front-end demethanizer scheme.



Front-end deethanizer scheme.

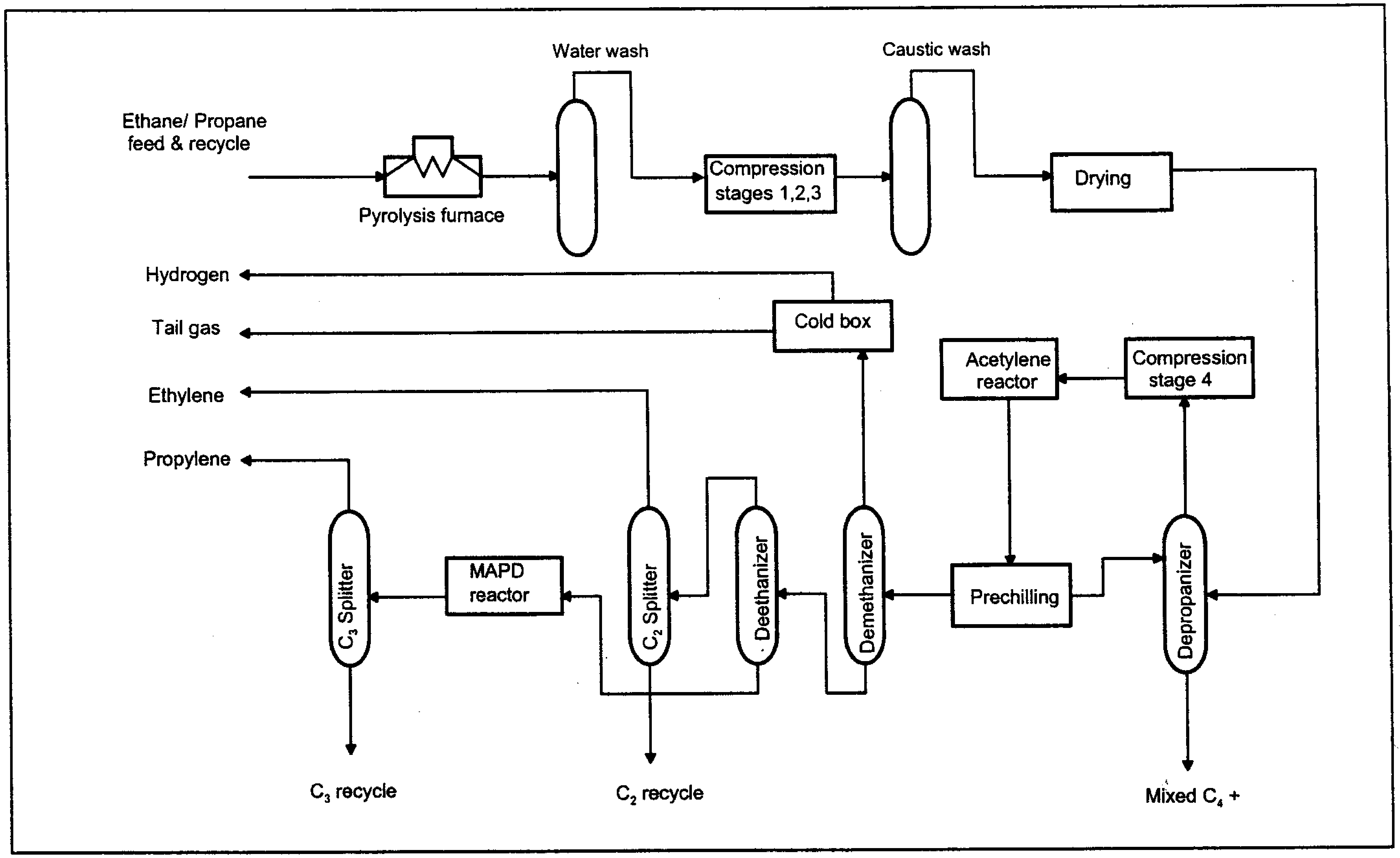
3-5 (continued)

Front-end depropanizer

The cracked gas in this scheme shown on the next figure is compressed before being caustic-washed and dried. The dried cracked gas is sent to the depropanizer where the overhead is compressed. It then goes to a front-end acetylene reactor where the acetylene is hydrogenated to ethylene and ethane. After hydrogenation, the gases are partially condensed to provide reflux for the depropanizer. The net overhead liquid and vapor, which contain the C_3 and lighter components, go forward to the demethanizer prechillers. The net C_3 and lighter stream from the depropanizer are fractionated in the demethanizer, with the resulting bottoms stream being further fractionated in a deethanizer to a methyl acetylene/propadiene hydrogenation reactor. The reactor effluent is sent to the C_3 splitter.

In addition to the location of the demethanizer, deethanizer, and depropanizer, the location of the reactor for hydrogenating the C_2 and C_3 acetylenes is dictated by the scheme selected. In both the front-end deethanization and front-end depropanizer separation schemes, the C_2 acetylene hydrogenation converters are located ahead of the cracked gas chilling train and demethanizer. By contrast, in the front-end demethanization scheme, the acetylene converter follows the demethanizer and deethanizer. In this scheme, the hydrogen in the cracked gas has already been removed in the gas chilling train and demethanizer and an external addition of hydrogen is necessary in the acetylene converter. However, hydrogenation of the C_3 acetylenes in all schemes requires an external addition of hydrogen.

Separation of the ethylene-ethane mixture also provides several choices. This separation is commonly achieved in a high-pressure column, utilizing close-cycle propylene refrigeration. This process can also be carried out by low-pressure fractionation, using open-cycle ethylene refrigeration. Alternatively, the selection can also be achieved with absorption using highly selective solvents. The latter separation technique is seldom used. However, the low-pressure fractionation has received some support since the relative volatility between ethylene and ethane is higher and the separation is easier, resulting in a lower reflux ratio and fewer trays than in the



Front-end depropanizer scheme.

3-5 (continued)

high-pressure column. However, this fractionation scheme requires the use of ethylene refrigeration instead of propylene refrigeration. Both fractionation schemes have advantages and disadvantages. For example, the lower design pressure reduces the wall thickness of the column with a resultant cost saving. However, this cost reduction is offset by the higher material cost of equipment and piping due to the lower design temperature.

A good example of the front-end demethanizer scheme is the licensed processes developed by the M.W. Kellogg Company, *Hydrocarbon Process.* **68**(11), 102(1989), and by the Stone and Webster Engineering Company, *Hydrocarbon Process.*, **64**(11), 137(1985). The process developed by Linde A.G., *Hydrocarbon Process.* **64**(11), 137(1985) for producing ethylene from hydrocarbon liquids is an example of the deethanizer scheme. The advantages and disadvantages of these processes are listed below. (Other advantages and disadvantages of these processing schemes were listed earlier.)

Kellogg scheme advantage:

High olefin yields due to the very small (less than 0.1 second) reaction times in the pyrolysis furnace unit. Process handles feeds ranging from ethane to vacuum gas oils. Provides ultra-purified hydrogen stream for catalytic conversion. Generates excess high pressure steam for use elsewhere in a multi-purpose plant.

Kellogg scheme disadvantages:

Steam pyrolysis is performed at temperatures higher than most other ethylene processes creating more coking potential. Process requires two separate water quench exchangers.

Stone and Webster scheme advantages:

Handles feeds ranging from gases to liquid hydrocarbons. Process is characterized by a highly selective pyrolysis coil. Steam pyrolysis temperature not as high as the Kellogg process reducing

3-5 (continued)

the coking potential. Operating history of quench exchanger shows it to be basically non-fouling. Removes acid gas with patented amine process.

Stone and Webster scheme disadvantages:

Requires an oil quench (with liquid feed stocks) and also a water quench. Ethylene yields slightly less than that obtained with the Kellogg process.

Linde AG advantages:

Feed is cracked in more conventional tubular furnaces which can be adapted to meet specific requirements of high capacity, high severity, or high selectivity. Recompression of cracked gases is to a lower pressure than other processes.

Linde AG disadvantages:

Operates only on hydrocarbon liquids. Requires an oil and a water quench.

3-6

To assist in preparing the material balance for the process shown in Fig. 3-13, the information provided below can simplify the calculations. The flow of ethylene in stream 533 is 62,008 kg/h. From the preliminary design, the $C_4 - C_6$ component listed in streams 413 and 416 includes 1-3 butadiene, i-butene, 1-butene, n-butane, C_{5s}PON, benzene, and toluene. The calculated component distribution in kg/h is

Component	Stream 413		Stream 416	
acetylene			0.005	
ethylene			0.005	
ethane			1.2	
MAPD	97.5		458	
propylene	2,220		20,572	
propane	1912		14,321	
1-3 butadine	1635		2790.2	
i-butene	294.2		563	
1-butene	275.5		515	
n-butane	45.4		67.8	
C_{5s} PON	2250	8734	715	4776
benzene	3476		122.5	
toluene	757.9		2.5	
heavy gasoline	1858		0.55	

Design information:

Distillation columns were simulated and designed with the CHEMCAD-SCDS method using the Soave-Redlich-Kwong equation of state. Reflux ratio for C-601 was set at 1.5 R_{min}. For C-602, C-603, C-604, and C-605 it was 1.2 R_{min}. Cooling water was available with an inlet temperature of 29°C and an outlet temperature of 35°C. Plate efficiency of the valve trays was assumed to be 85%. Compressor K-601 assumes a polytropic efficiency of 85%. Removal of MAPD (methylacetylene and propadiene) assumes a level of the C_3 acetylene at the exit of 2500 ppm in reactor R-60 and an exit level less than 50 ppm in reactor R-602. The Arrhenius equation parameters for the two C_3 acetylene removal and propylene hydrogenation reactions specified as

3-6 (continued)

$$C_3H_4 + H_2 \rightarrow C_3H_6 \qquad (1)$$

$$C_3H_6 + H_2 \rightarrow C_3H_8 \qquad (2)$$

are

	Activation energy, E cal/g mol	Temp range, °C	Frequency factor, A
(1)	1.18×10^4	100 – 130	2.3×10^6
(2)	1.17×10^4	100 - 130	5.25×10^3

The CC-Therm software was used for the thermal and mechanical design of all shell-and-tube heat exchangers including E-601 through E-612.

The overall mass balance was developed with the use of ASPEN PLUS and CHEMCAD-III software programs utilizing the Soave-Redlich-Kwong equation of state relation. This equation of state provides a good match between simulated properties and actual properties reported in the literature.

To complete the material balance, components in streams 215 and 325 are provided in kg/h as

Component	Stream 215 (77.3°C, 179 kPa)	Stream 325 (62.8°C, 234 kPa)
$C_{5,s}$ PON		1.8
benzene		25.2
toluene		25.2
heavy gasoline	9.1	483.1

Figure 3-15 provides a summary of the mass balances around various parts of the separation section and can be used to verify results.

The material balance result is summarized in the following table with each component flow provided in kg/h.

3-6 (continued)

Material balance results in kg/h for the product separation section shown in Fig. 3-13.

	Stream numbers							
Components	602, 621	603	604	605,605	609	610,611	612,613	614
hydrogen					27.8	27.8		16.3
methane	2.9							
acetylene	0.02	0.22	0.005			0.005	0.005	
ethylene	61,946	62.5	0.005			0.005	0.005	
ethane	59.9	11,893	1.2			1.2	1.2	
MAPD[†]			555	0.6		555	99.9	
propylene		2.1	22,791.5	0.06		22,791.5	23,166.7	
propane			16,232.3	0.5		16,232.3	16,340.2	
1,3 butadiene			1.3	4423.6		1.3	1.3	
i-butene			0.6	856.2		0.6	0.6	
1-butene			0.4	790.1		0.4	0.4	
n-butane				113.2				
$C_{5,s}$PON[‡]				2964.8				
benzene				3598.6				
boluene				760.4				
heavy gasoline				1859				

	Stream numbers					
Components	615	616,617,618	619	620	607	608
hydrogen	16.3	0.01	0.01			
methane						
acetylene	0.005	0.005	0.005			
ethylene	0.005	0.005	0.005			
ethane	1.2	1.2	1.2			
MAPD[†]	99.9	1.9		1.9	0.6	
propylene	23,166.7	23,031.6	23,004.8	26.8	0.07	
propane	16,340.2	16,589.4	94.8	16,494.6	0.5	
1,3 butadiene	1.3	1.3		1.3	4423.5	0.05
i-butene	0.6	0.6		0.6	856.2	
1-butene	0.4	0.4		0.4	790.1	0.005
n-butane					113.2	0.01
$C_{5,s}$PON[‡]					0.03	2964.8
benzene						3598.6
boluene						760.4
heavy gasoline						1859.0

[†] Methyl acetylene/propadiene

[‡] Paraffin, olefin, napthene

3-7

The material balance from Problem 3-6 and either ASPEN PLUS or CHEMCAD-III computer software is used to develop the energy balance around each piece of equipment in the ethylene separation section. For example, around distillation column, C-601, the computer program establishes the heat content of streams 533, 602, and 603 above a selected datum plane. The distillation calculation indicates the flow rates of the overhead and bottoms streams. The reflux and reboil then indicate the flow rates of the streams that are returned to the column and permits evaluation of the condenser and reboiler duties. In kW this can be expressed as

$$h_{533} + \dot{q}_{E-602} - (\dot{q}_{E-601} + h_{602} + h_{603}) = 0$$

$$15{,}208 + 19{,}479 - (15{,}639 + 29{,}674 - 10{,}626) = 0$$

Thus $\dot{q}_{E-601} = -15{,}639\,\text{kW}$

$$\dot{q}_{E-602} = 19{,}479\,\text{kW}$$

Similarly

$\dot{q}_{E-603} = -4422\,\text{kW}$

$\dot{q}_{E-604} = 8050\,\text{kW}$

$\dot{q}_{E-605} = 1245\,\text{kW}$

$\dot{q}_{E-606} = -636\,\text{kW}$

$\dot{q}_{E-607} = -325\,\text{kW}$

$\dot{q}_{E-608} = -30{,}752\,\text{kW}$

$\dot{q}_{E-609} = 29{,}288\,\text{kW}$

$\dot{q}_{E-610} = -1432\,\text{kW}$

$\dot{q}_{E-611} = 890\,\text{kW}$

$\dot{q}_{E-612} = -507\,\text{kW}$

Equipment dimensions have already been summarized in Table 3-7, p. 111. These dimensions are to be verified in the problem.

3-8

The material balance for Fig. P3-8 was developed with the use of ASPEN PLUS and CHEMCAD-III software programs utilizing the Soave-Redlich-Kwong (SRK) equation of state relation. This equation of state provides a good match between simulated properties and actual properties reported in the literature.

The distillation column was simulated and designed with the CHEMCAD-SCDS method, also using the SRK equation of state. Reflux ratio for the distillation column was set at 1.2 R_{min}. Plate efficiency of the valve trays was assumed to be 0.85. Cooling water was available with an inlet temperature of 29°C and an exit temperature of 35°C. The CC-Therm software was used for the thermal and mechanical design of all the shell-and-tube heat exchangers.

The material balance in kg/h for various locations in Fig. P3-8 are given below

	Stream no.			
	1, 2, 3, 4	5	6	7, 15, 16
Ethylene, kg/h	62,014	29.4	61,986	55,352
Ethane, kg/h	11,956	5776.6	6178	60
	73,970	5806	68,164	55,412
Ethylene, mol fract.	0.8475	0.0055	0.915	0.999

	Stream no.			
	8, 9	10	11, 12, 13, 14	17, 18
Ethylene, kg/h	6632.5	6597.2	35.3	61,949.2
Ethane, kg/h	6119.5	0.8	6118.7	60.8
	12752	6598	6154	62,010
Ethylene, mol fract.	0.5373	0.999	0.0055	0.999
		Answer		

3-8 (continued)

An energy balance around the pieces of equipment in Fig. P3-8 provides the heat duties for the seven heat exchangers, condenser and reboiler as:

E-S1, E-52 $\dot{q}_{2-3} = 4053\,\text{kW}$

E-S3 $\dot{q}_{3-4} = 2671\,\text{kW}$

E-S4 $\dot{q}_{8-9} = -80\,\text{kW}$

E-S4 $\dot{q}_{12-13} = 80\,\text{kW}$

E-S5 $\dot{q}_{\text{cond}} = -3005\,\text{kW}$

E-S6 $\dot{q}_{\text{reboil}} = 2000\,\text{kW}$

E-S7 $\dot{q}_{11-12} = 770\,\text{kW}$

E-S8 $\dot{q}_{15-16} = -2384\,\text{kW}$

E-S9 $\dot{q}_{17-18} = -2205\,\text{kW}$

Answer

The enthalpy content in kW of each stream based on a fixed datum plane is given below:

Stream no.	Enthalpy content	Stream no.	Enthalpy content
1, 2	15,207 kW	10	3139
3	19,260	11	-5671
4	21,931	12	-4901
5	-4542	13, 14	-4821
6	26,474	15	31,006
7	27,938	16	28,622
8	-1466	17	33,856
9	-1546	18	31,651

Answer

3-9

Without information on the process streams exchanging heat in E-S1, E-S2, E-S3 and E-S7, the areas required for these heat exchangers cannot be evaluated. CC-Therm results for these four exchangers and their cost using Fig. 14-18 are as follows:

Exchanger data	E-S1	E-S2	E-S3	E-S7
Heat duty, kW	3271	782	2671	770
$\Delta T_{\ln m}$, °C	12.2	26.2	9.8	18.6
U, $W/m^2 \cdot K$	756	310	636	312
A, m^2	355	96	429	133
Cost, $ (Jan. 2002)	27,500	11,800	31,000	14,500

For the other heat exchangers, condenser, and reboiler, the sizes of the heat exchangers can be evaluated to provide

Exchanger	Area, m^2	Cost, $ (Jan. 2002)
E-S4, $\dot{q}_{8\text{-}9}$	68	9,400
E-S5, $\dot{q}_{\text{cond}}$	449	32,800
E-S6, $\dot{q}_{\text{reboil}}$	98	11,900
E-S8, $\dot{q}_{15\text{-}16}$	443†	32,000
E-S9, $\dot{q}_{17\text{-}18}$	488†	34,000

† Utilizes cooling water with an inlet of 29°C and an outlet of 35°C. With the CHEMCAD-SCDS computer software, the information for the distillate column, assuming 85% tray efficiency, is

Diameter, 1.52 m
Number of actual trays, 57
Tray spacing, 0.61 m
Installed cost from Fig. 15-15 ($1250/tray), $71,250
(For purchased cost, approximately $57,000)

The power required by the two centrifugal rotary compressors and their cost from Fig. 12-28 are

Compressor (7 to 15)	3008 kW	$700,000
Compressor (10 + 16 to 17)	2722 kW	$650,000

3-9 (continued)

The area and cost[†] of the two membrane units are

Membrane (4 to 5 and 6)	8900 m^2	$2,314,000
Membrane (6 to 7 and 8)	995 m^2	$259,000

[†] Cost of membranes is \$260/m^2 based on vendor quotation.

Answer

3-10

A membrane cascade system with recycling and utilizing facilitated transport technology can be shown to recover ethylene from ethane at the required polymer-grade composition. The separation can be achieved without the aid of a distillation column. Two or more membrane units with associated compression equipment are sufficient for separating ethylene from ethane at high purity and recovery. However, the high recovery rate of ethylene as well as the high purity product dictates high recycle rates between membrane units and large membrane area requirements.

Figure P3-10 shows a two-stage membrane cascade with recycling for producing ethylene product with a polymer-grade composition of 0.999 mol percent. The fresh feed pressure of 2605 kPa and temperature of 28°C are identical to the conditions shown for stream 4 in the series configuration hybrid system presented in Fig. P3-8. The process conditions for this cascade configuration are shown in the accompanying table.

Process conditions for the two-stage membrane system in Fig. P3-10

	Stream no.					
	1	2	3	4	5	6
Temp., °C	28	33	31	31	10.9	43.2
Press., kPa	2605	2605	2605	2605	827	1344
Ethylene, kg/h	62,014	91,072	153,086	64.8	153,021.2	153,021.2
Ethane, kg/h	11,956	17,558	29,514	11,895.2	17,618.8	17,618.8
Ethylene mol fract.	0.8475	0.8475	0.8475	0.0056	0.9029	0.9029

	Stream no.					
	7	8	9	10	11	12
Temp., °C	33	33	80.4	20.2	237	33
Press., kPa	1337	1337	2619	103	1970	1944
Ethylene, kg/h	153,021.2	91,072	91,072	61,949.2	61,949.2	61,949.2
Ethane, kg/h	17,618.8	17,558	17,558	60.8	60.8	60.8
Ethylene mol fract.	0.9029	0.8475	0.8475	0.9990	0.9990	0.9990

3.30

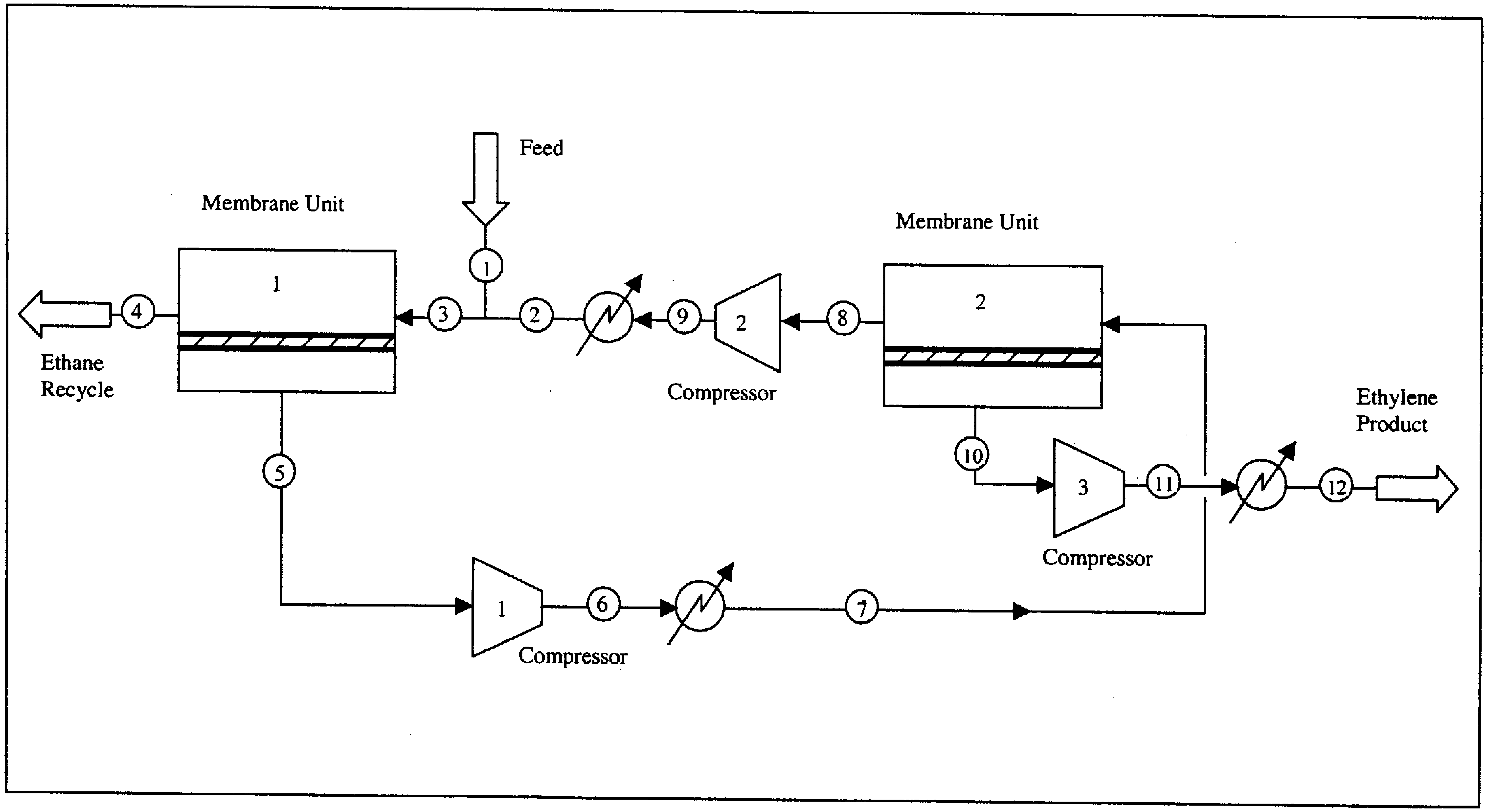


Figure P3-10 : Membrane cascade for ethane/ethylene separation.

3-10 (continued)

In this membrane cascade system, the fresh feed stream 1 is mixed with the recycle in stream 2 and then fed to the first membrane stage. In the first membrane unit, the recycled ethane is separated as the retentate and the recovered ethylene is concentrated as the permeate. There is a trade off between the membrane area requirement and the operating pressures. Analysis shows that the optimal permeate pressure for the first membrane is close to 827 kPa. This pressure balances the trade off between the compression costs and the capital costs. The overall performance of the first membrane stage is lower when compared to the second membrane stage. This low performance is because of the high percentage of ethylene recovered in the permeate side as well as the lower pressure ratio of the first stage which requires a very large membrane area.

The permeate stream after leaving the first membrane stage is compressed to 1337 kPa and cooled to 33°C before being sent to the second membrane stage. This stream could be compressed to the pressure of the fresh feed in a two-stage compressor before being sent to the second membrane stage. However, the reduction in compression load due to the reduction in flow rate to the second stage compressor can be shown to be more significant than theeffect the lower pressure would have on the membrane area requirement.

Polymer-grade ethylene is produced in the second membrane stage at a permeate pressure of 103 kPa. Permeate pressure can be increased to lower the ethylene product compression cost, but this can only be achieved when the membrane feed composition is above its minimum value. When the permeate pressure is increased, the required feed composition is increased. Also, the membrane area required increases as the permeate pressure is increased. Finally, an increase in permeate pressure requires an increase in the recycling rate with an increase in compression costs. The second-stage performance is more efficient than the first stage with respect to both the selectivity and flux rate. This performance is affected by the feed and permeate pressures as well as by the feed, retentate, and permeate compositions.

3-10 (continued)

Even though the pressure difference has a significant impact on the membrane separation process, the pressure ratio is more important than the pressure difference for this particular process. Higher pressure ratios increase both the selectivity and flux rate and these lead to improved membrane unit performance.

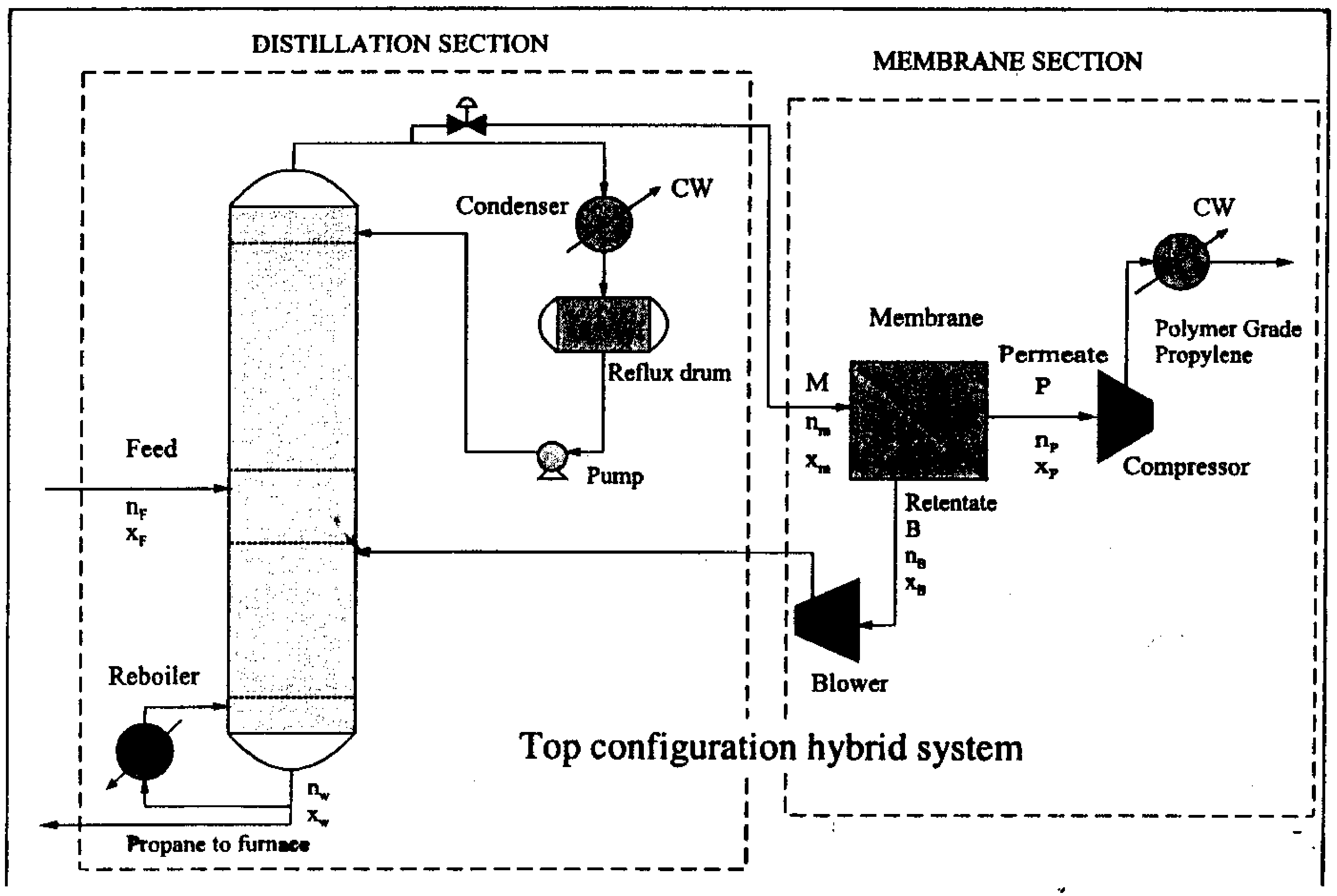
3-11

The separation of propylene from propane utilizing facilitated transport technology is similar to that used in the separation of ethylene from ethane. Figure 3-6 provides several possible configurations for the hybrid distillation membrane separation system. Another possible configuration is the combination of the separate top and bottom configurations into a hybrid system that utilizes a membrane separation system at both ends of the distillation column. Schematics of these different hybrid systems are provided in the figures below.

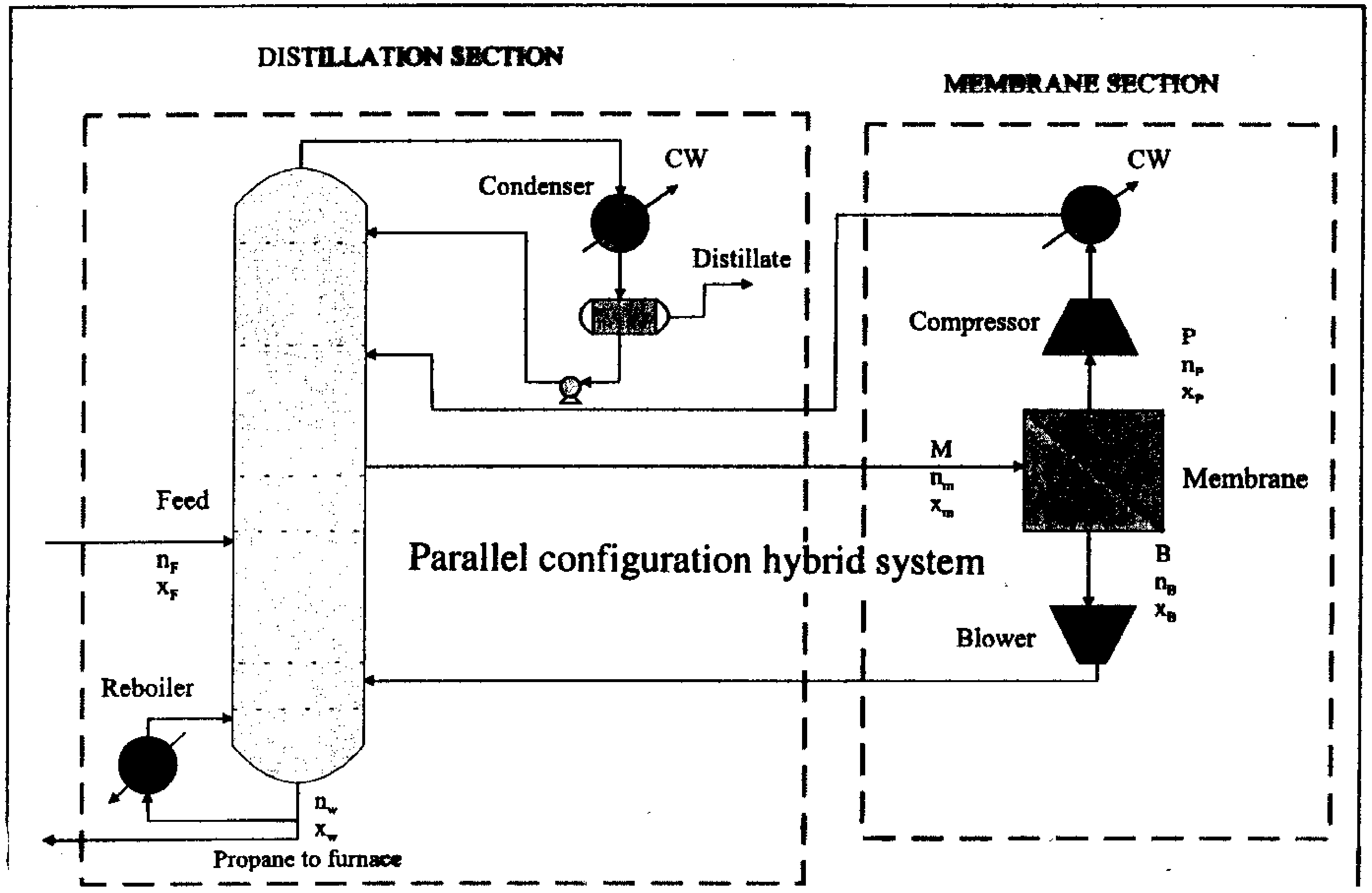
A review of these four different configurations indicates that the top configuration will provide a reduction of separation stages in the rectification section of the distillation column. This reduction in the number of stages will only have a small effect on process and energy savings over the conventional C_3 splitters. The parallel configuration will have less effect on the number of stages required, but should provide greater energy savings since both ends of the column receive benefit from this configuration. The bottom configuration can also provide a greater reduction in the stages required for separation and will have an energy savings in the reboiler. The top-bottom configuration offers the best opportunity to obtain both a capital savings and an energy savings. Thus, this configuration appears to offer the greater annual savings and warrants further evaluation.

Answer

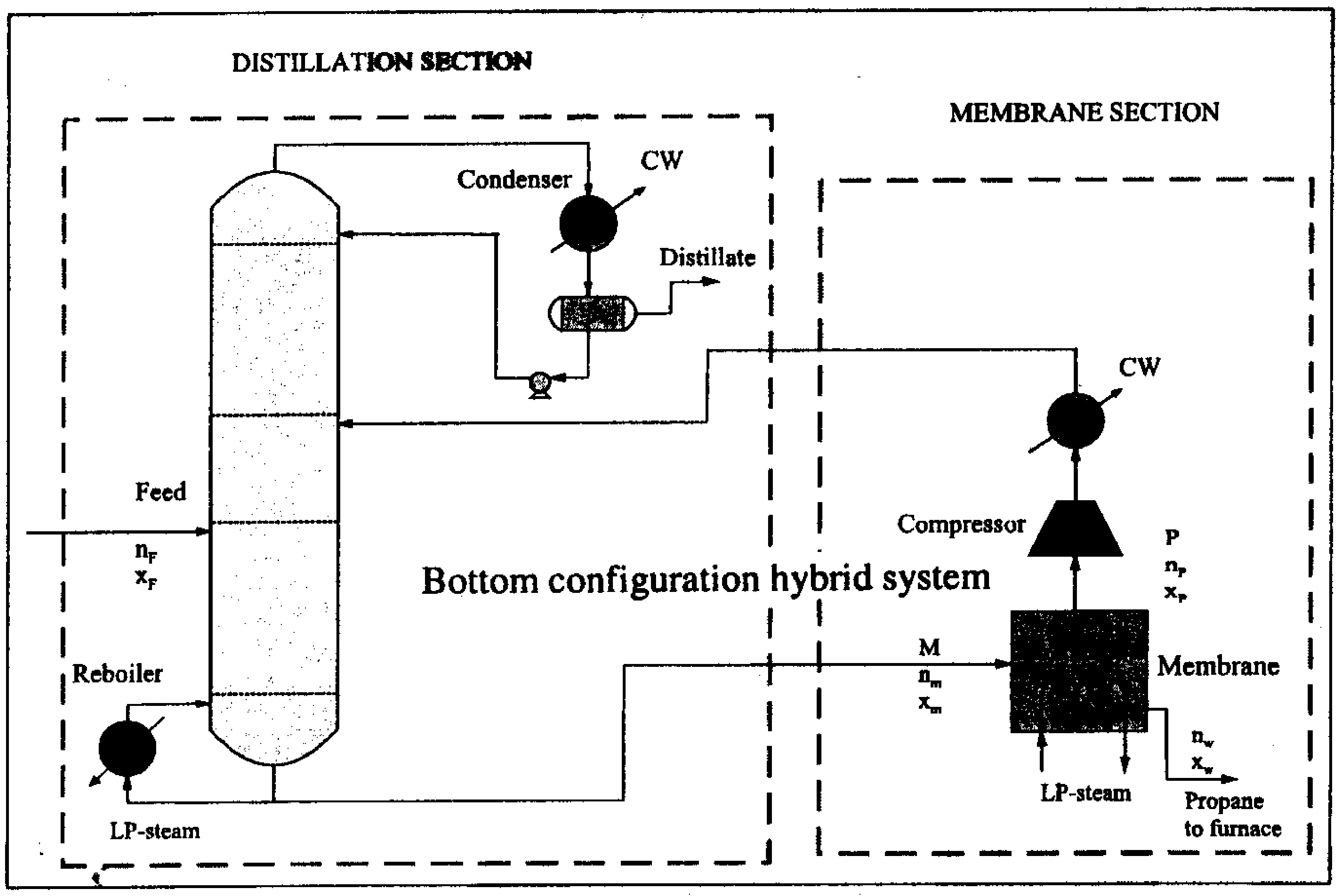
Note: the four configurations have been investigated and the top-bottom configuration provided the largest capital and processing savings among these configurations.



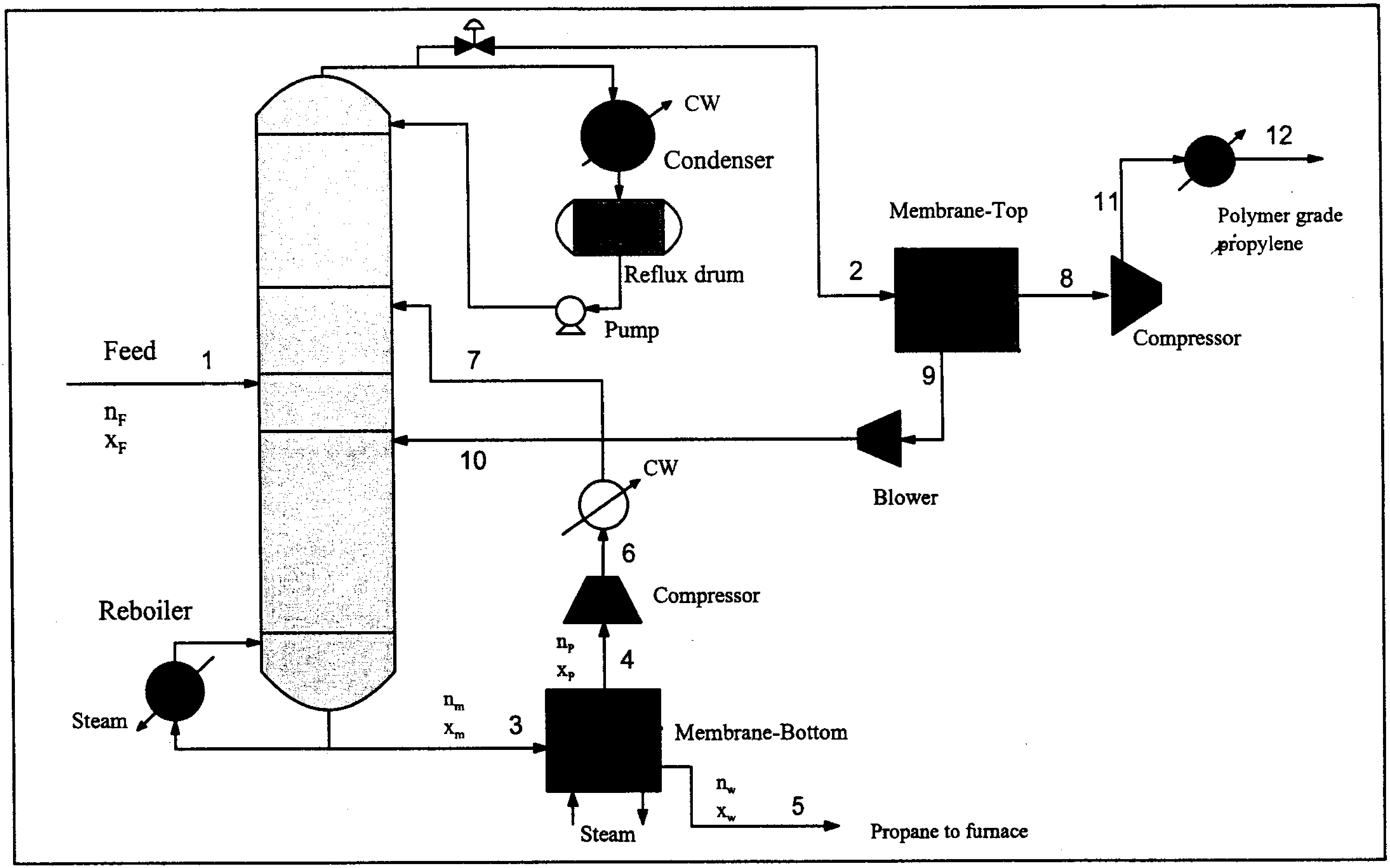
Top configuration hybrid system



Parallel configuration hybrid system



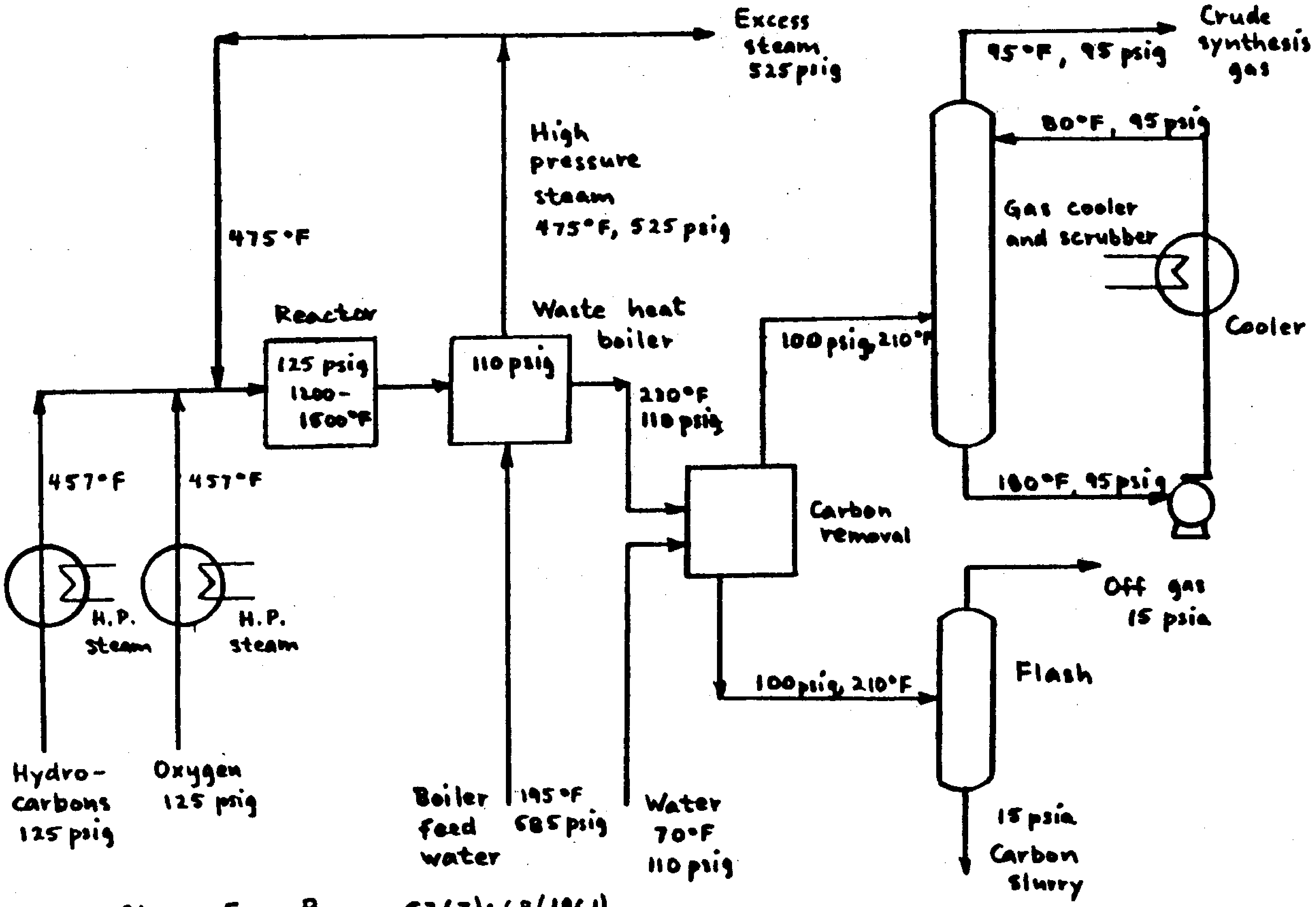
Bottom configuration hybrid system



Top-bottom configuration hybrid system (*TBCHS*) for the propane/propylene separation.

3-12

A process flow sheet for preparing synthesis gas from any hydrocarbon is given below.

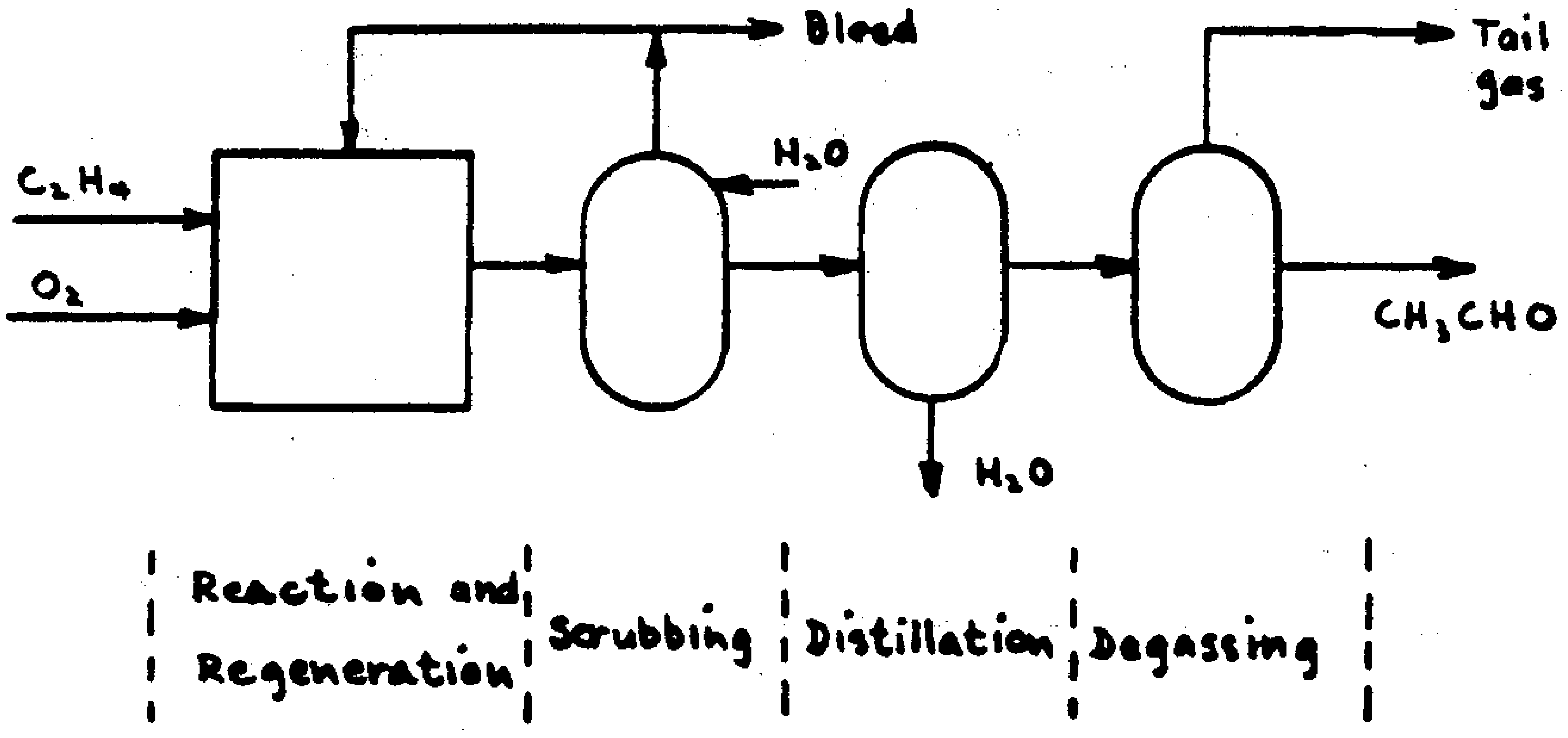


see Chem. Eng. Progr., 57(7): 68(1961)

The operating conditions for the synthesis gas process using the given feed stock, a heavy fuel oil, are presented in an article by S.C. Singer and L.W. ter Haar in *Chem. Eng. Progr.*, 57(7), 68(1961). The material balance may be made either with software programs by ASPEN PLUS or CHEMCAD and using these operating conditions and the final composition of the crude synthesis gas.

3-13

The simplified flow sequence for producing acetaldehyde from ethylene is given below:



Reaction and regeneration:

The main considerations are the reaction kinetics, equilibrium relations, and heat of reaction. The relationship of these factors will determine the reactor configuration and the cooling water requirements. Obviously, physical and thermal data for all the flow streams will be needed. Catalyst activity with information on its decline with time and reactivation time will be needed for regeneration design purposes.

Scrubbing:

The equilibrium relationship between the various reactor components and water will be the important design factor in this unit. Physical and thermal data for all the flow streams will be needed. Build up of inerts is controlled by a bleed stream. The composition of the latter must be known to determine the requirements for an auxiliary reactor designed to complete the oxidation of unconverted ethylene.

3-13 (continued)

Distillation:

The distillation unit requires equilibrium and vapor pressure data for the mass transfer calculations. Separation of key components must be specified and separation efficiencies need to be established. Physical and thermal data for all the flow streams will again be needed. Heat transfer coefficients will be required for the design of heat exchangers, condenser, and reboiler.

Degassing:

This unit will require much of the same information that is required for the distillation unit only with different compositions and flows.

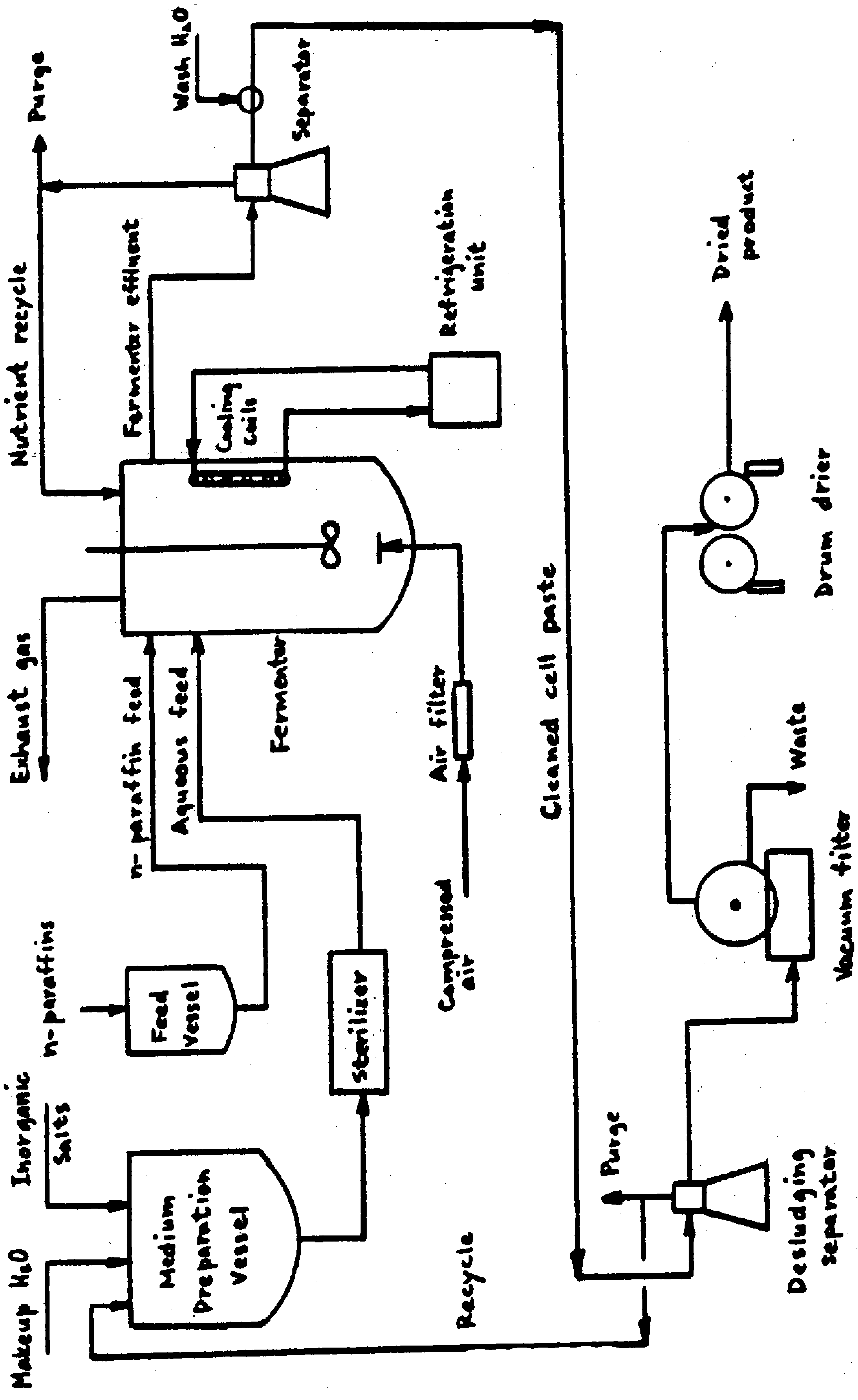
A simplified equipment flow sheet for the acetaldehyde process is given in *Hydrocarbon Process.* **44**(11), 159(1965). Since this is a proprietary process, only sketchy information is supplied with regards to temperature, pressure, and stream composition at each piece of equipment. Some of this information can be deduced from the information supplied for the raw materials and utilities requirements per short ton of acetaldehyde.

3-14

A simplified equipment flow sheet for the acetaldehyde process is given in *Hydrocarbon Process.* **46**(11), 135(1967). In this variation of the process, the spent catalyst is reoxidized with air in a separate regenerator. Ethylene is converted to acetaldehyde in a single-pass contact with the catalyst solution at a slightly higher pressure than used in a single-stage unit. In a separator following the reactor, the pressure is reduced and reaction heat vaporizes the acetaldehyde product from the catalyst solution. The spent catalyst from the separator is pumped to the regenerator where it is reoxidized by contact with air. A comparison of raw materials and utilities required for the two-stage acetaldehyde process may be made with that given for an actual 75,000 short ton/year plant discussed in the same reference noted above. Since the solution of this problem is quite long, it is not presented here.

3-15

A flow sheet outlining the conversion of petroleum to food supplements using organic microorganisms is presented below;



3-15 (continued)

In the process an aqueous feed containing the necessary inorganic nutrients is mixed in a vessel before being sent to the fermentor with a metered amount of paraffinie feed stock. Compressed and sterilized air serves as the oxygen supply for the growing cells in the fermentor.

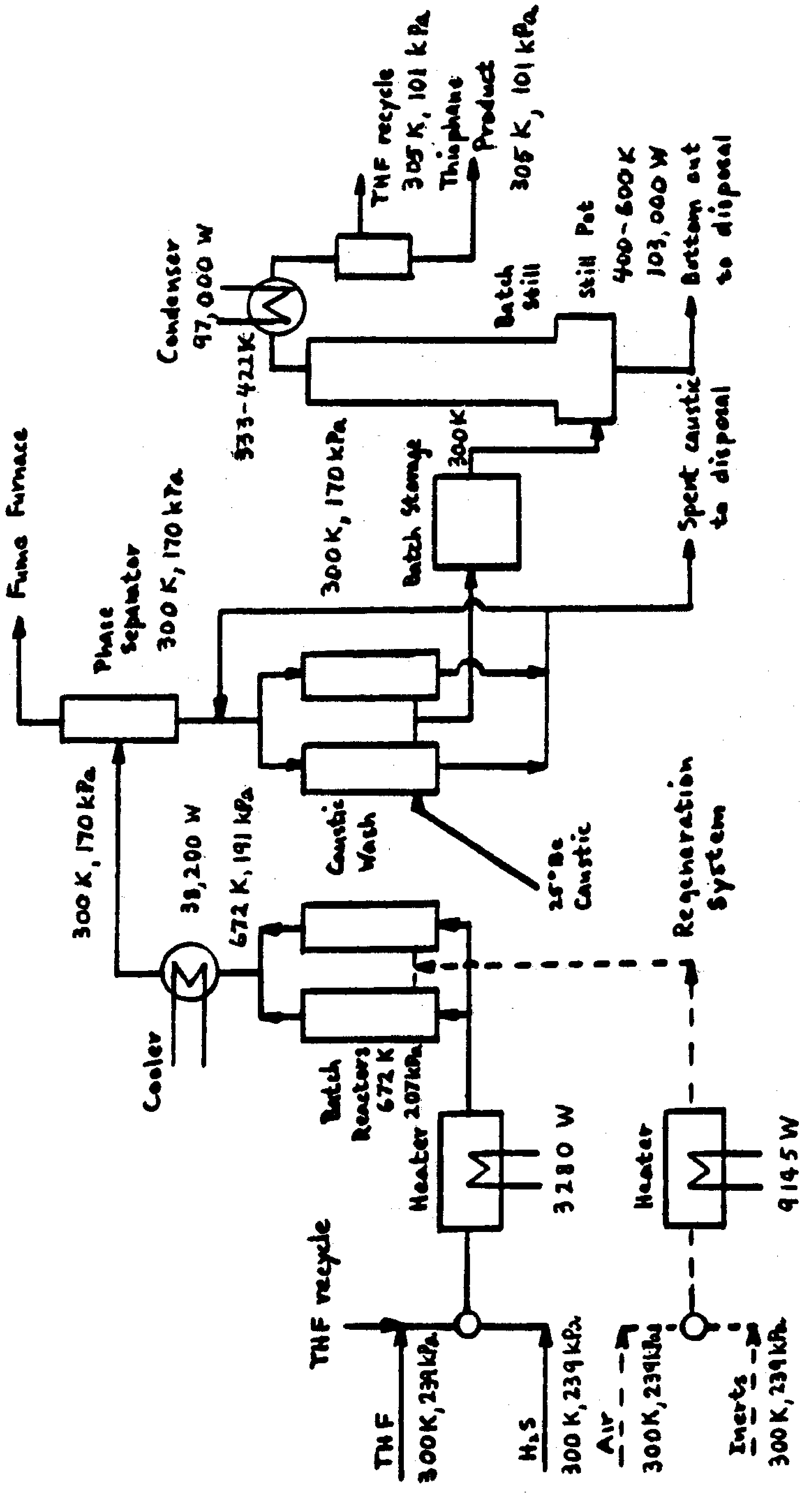
The growth of single cells is an autolytic process. The exponential growth rate can be maintained as long as the concentrations of all nutrients are kept above their respective critical values in the fermenting brew. The ratio of the paraffins to aqueous feedrates need to be maintained as high as possible. The temperature during fermentation must be kept constant by the careful control of heat removal.

At steady state, the product stream is continuously removed from the fermentor at a rate equal to feedrate. The product stream is concentrated in a desludging separator to a paste of 20 to 30% dry solids. After washing with warm water to remove adsorbed or absorbed hydrocarbons, a second desludging separator produces a cleaned paste that is further dehydrated in a rotary vacuum filter. A final drying step in an atmospheric drum dryer reduces the moisture content to less than 10%, and the powdered product is packaged for the customer.

The potential of using crude oil to produce human or animal-feed supplements has been recognized since the early 1960s. Many of the problems to accomplish this feat were listed by Wang in a *Chem. Eng.* 75(18), 99 (1968) publication. Several of these problems include separation of straight-chain hydrocarbons from isoparaffins, cyclic paraffins, and napthenes, development of a continuous non-aseptic processing technique, better understanding of the mass transfer characteristics of the process, control of the heat of fermentation, establishment of the required cell separation and purification technique, and improvement of the overall economics of the system. A number of these problems have been addressed over the past 35 years and suitable solutions have been formulated. However, the high demand for crude oil in meeting the world's transportation needs has increased the cost for crude oil to such a level that it is impractical to consider the use of this commodity to produce human or animal-feed supplements.

3-16

The process flow diagram for the production of 1.40 x 10^{-2} kg/s of thiophane is presented below with appropriate temperature, pressure, and heat exchanger duties. The material balance can be developed with the aid of suitable computer software based on the conditions listed on the flow sheet.

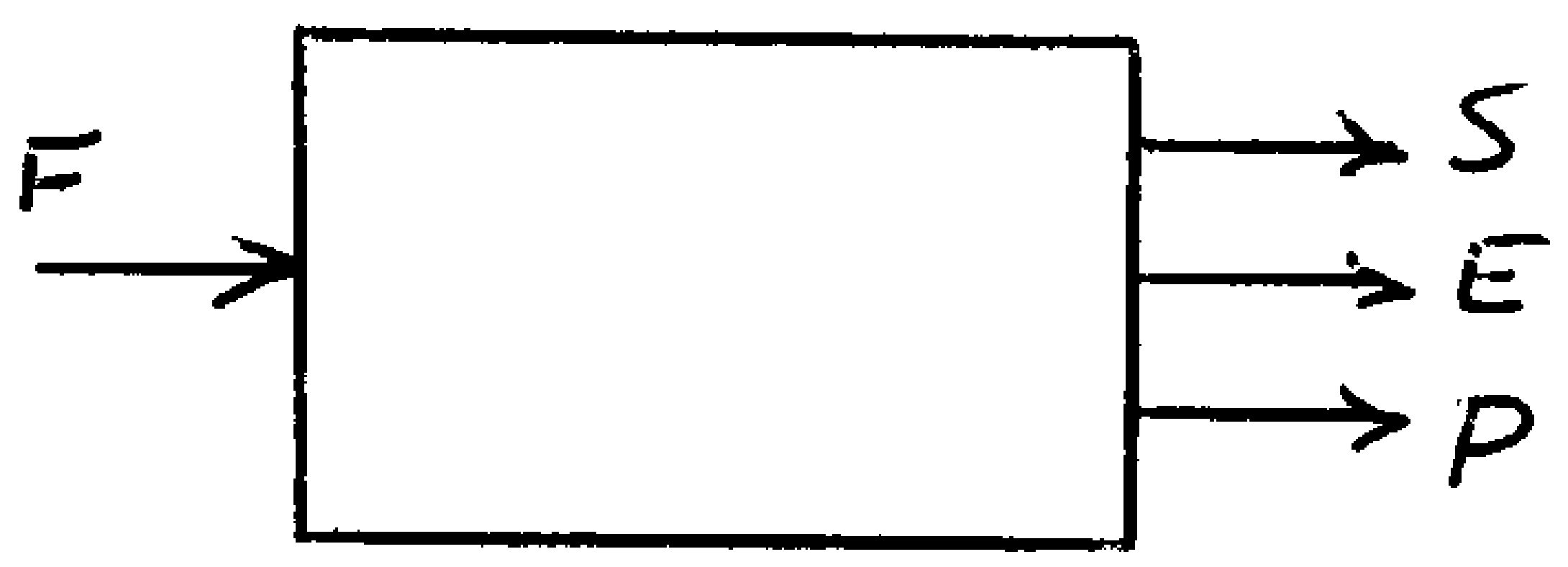


PROBLEM 4-1

Basis: 1 hour, 2.5×10^3 kgmol feed

Feed contains 83, 12, 4 and 1 mol % of methane, ethane, propane and n-butane, respectively. Let F, S, E and P represent the flowrates of feed, sales gas, ethane product and propane product, respectively.

a) I/O



b) Mole balances

Methane $(0.83)F = (0.995)S + (0.01)E$

Ethane $(0.12)F = (0.005)S + (0.97)E + (0.02)P$

Total $F = S + E + P$

Solving gives, $\underline{S} = (0.8330)F = \underline{2.083\times10^3\ \text{kgmol/h}}$

$\underline{E} = (0.1184)F = \underline{0.296\times10^3\ \text{kgmol/h}}$

$\underline{P} = (0.0486)F = \underline{0.121\times10^3\ \text{kgmol/h}}$

c) $LHV = \sum (\text{mol fr.})_i (LHV)_i$

Feed $LHV = (0.83)(0.802) + (1.428)(0.12) + (2.044)(0.04) + (2.686)(0.01) = 0.946\ \text{GJ/kg mol}$

Sales Gas $LHV = (0.995)(0.802) + (0.005)(1.428) = 0.805\ \text{GJ/kg mol}$

$MW_{ave} = \sum (\text{mol fr.})_i (MW)_i$

Ethane Prod. $MW = (0.01)(16) + (0.97)(30) + (0.02)(44) = 30.14$

Feed value $= (2.5\times10^3)(4.00)(0.946) = \underline{\underline{\$9.46\times10^3/h}}$

ANS.

Problem 4-1 (continued-1)

Sales gas value $= (2.083)(3.25)(0.805)$

$= \underline{\underline{\$5,450/h}}$ ANS

Ethane product value $= (0.296 \times 10^3)(0.40)(30.14)$

$= \underline{\underline{\$3,568/h}}$ ANS.

Propane product contains all butane in the feed, 0.02 mol fr. ethane, no methane, and the propane not included in other streams.

mols butane/h $= (0.01)(2.5 \times 10^3) = 25$

mols propane/h $= (0.04)(2.5 \times 10^3)$

$- (0.02)(0.296 \times 10^3) = 94.08 \times 10^3$

mols ethane $= (.02)((25 + 94.08)/0.98)$

$= 2.43$

Propane product value $= (0.44)(2.43 * 30$

$+ 94.08 * 44 + 25 * 58) = \underline{\underline{\$2459/h}}$

ANS.

d) $0 = (5,450 + 3,568 + 2,459) - 9,460 - C$

$C = 11,477 - 9,460 = \underline{\underline{\$2,017/h}}$

ANS.

C = Separation cost.

PROBLEM 4-2

NGL (natural gas liquids) consists of ethane, propane, iso- and normal butane, and C_5 and heavier hydrocarbons. See Chapter 3, Fig. 3-1 for information.

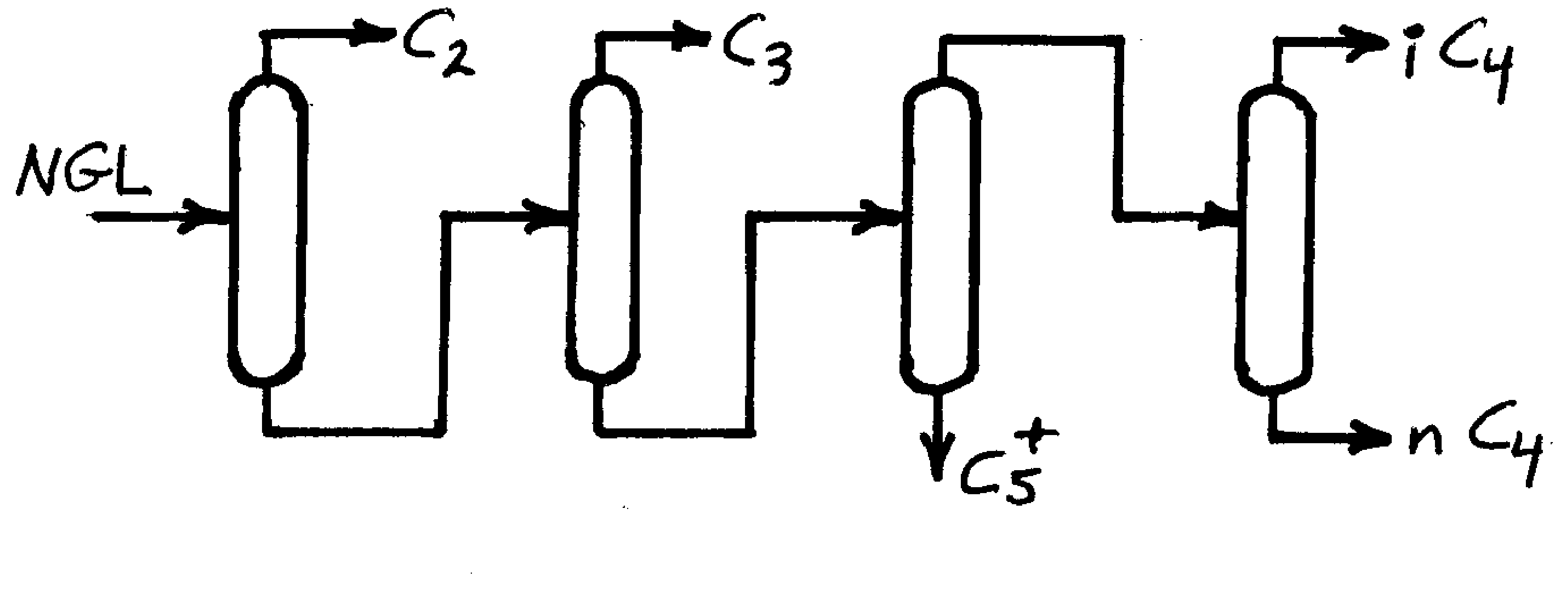
Pertinent properties are:

	Boiling pt. °C 101 kPa	506 kPa	1.01 MPa	Typical relative amt.
Ethane - C_2	-88.6	-52.8	-32	Most
Propane - C_3	-42.1	-1.4	26.9	2nd most
i-butane - iC_4	-11.7	39.0	66.8	4th most
n-butane - nC_4	-0.5	50.0	79.5	3rd most
gasoline - C_5^+		greater		least

Hueristics:

1) Remove most volatile as overhead products one-by-one
2) Remove largest quantities early
3) Save most difficult separation until last.

1) and 2) lead to removing C_2 in first col., C_3 in second col. 3) leads to separating (iC_4 + nC_4) from C_5^+, then separating C_4's. So, suggested train is

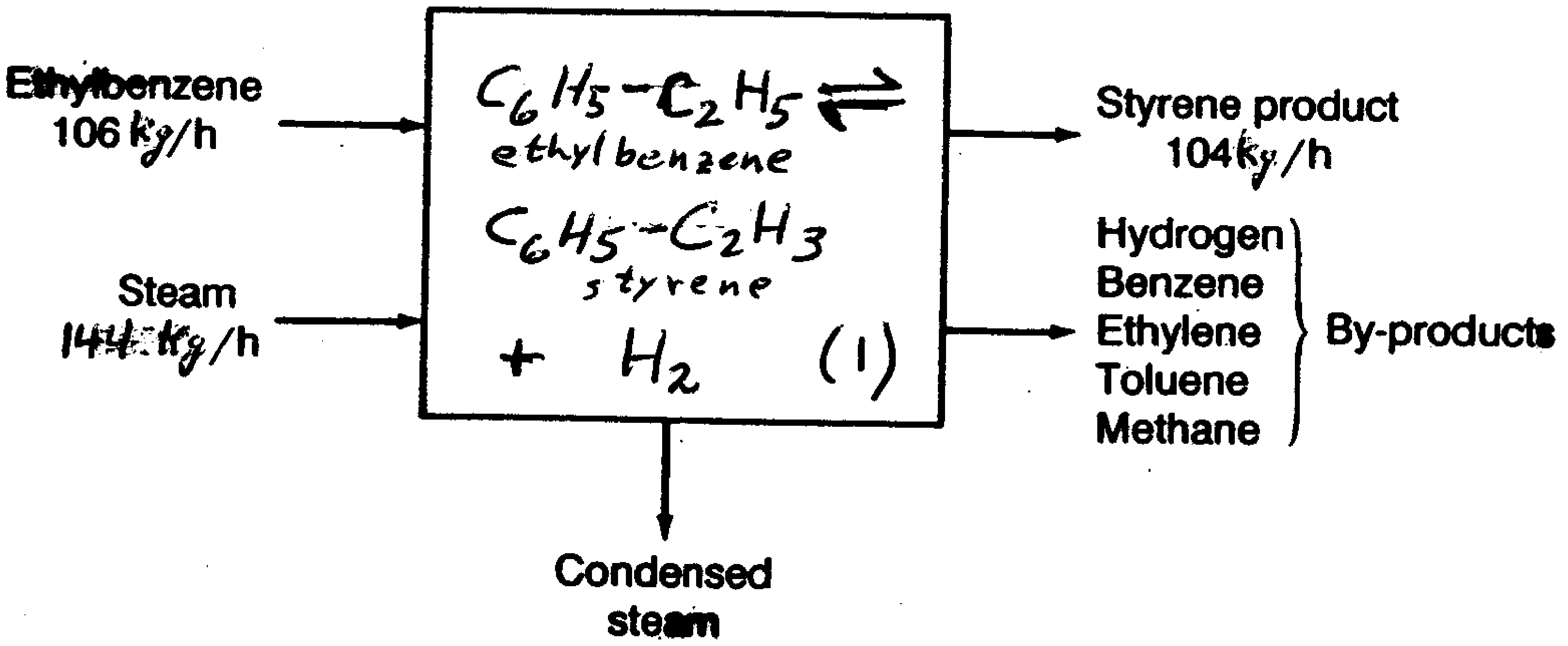


PROBLEM 4-3

Sources: Kirk-Othmer Encyclopedia of Chemical Technology, 4th ed., Vol. 22. Ullman's Encyclopedia of Industrial Chemistry, 5th ed, Vol. A25. Peters and Timmerhaus 4th ed. www.chemicalmarketreporter.com.

Styrene is made from ethylbenzene. The main process is catalytic dehydrogenation of ethylbenzene.

I/O



Values of \$0.57/kg for styrene, \$0.48/kg for ethylbenzene, and \$0.15/kg for hydrogen are used (the hydrogen value is based on its heat of combustion and a heat value of \$1.26/GJ). 1 kg-mol each of styrene (104 kg) and hydrogen (2 kg) requires 1 kg-mol of ethylbenzene (106 kg); Thus, on a basis of 1 kg-mol of styrene, the value of the products is

$$104 * 0.57 + 2 * 0.15 = \$59.6$$

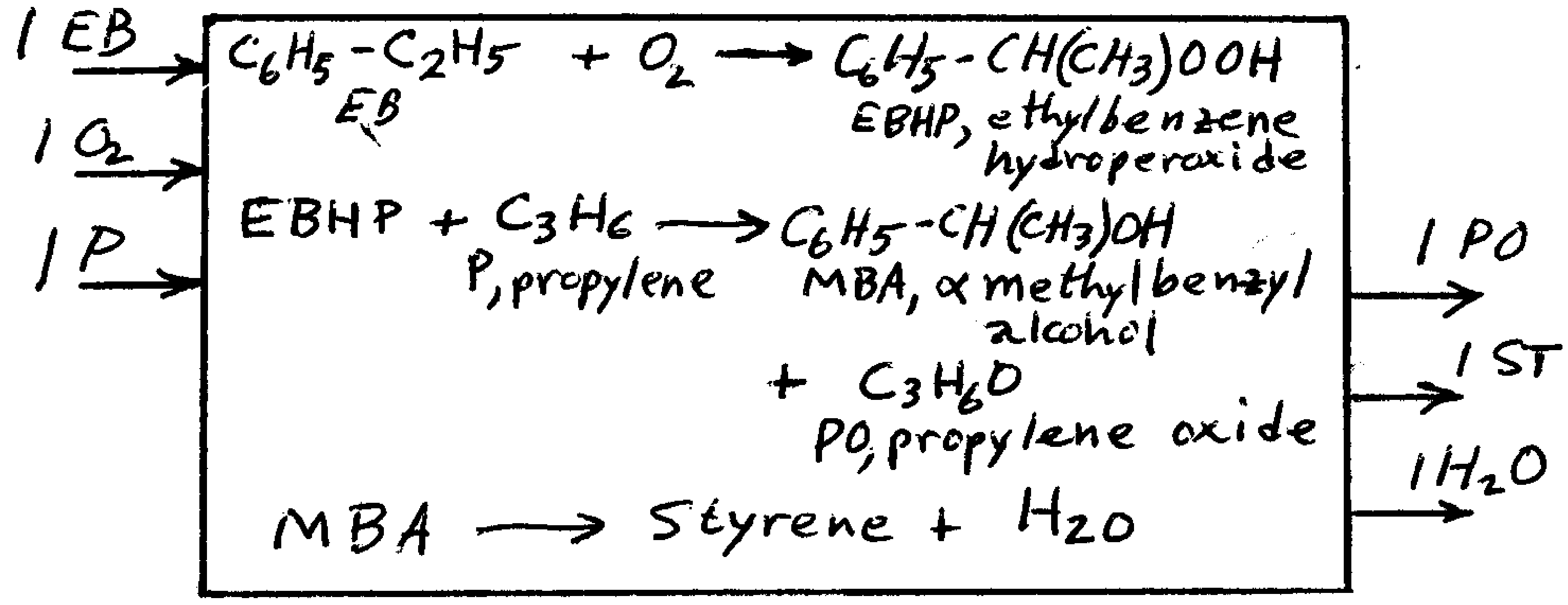
while the value of the ethylbenzene is

$$106 * 0.48 = \$50.1$$

The product value exceeds that of the raw materials, so it is possible for the process to be profitable and the process synthesis may continue. The cost of carrying out the process—the equipment, utilities, labor, overhead, etc.—has not been accounted for, nor has the fact that the conversion of reactant to product necessarily will be less than 100 percent.

PROBLEM 4-3 (continued -1)

Another commercial but less used reaction path to styrene (ST) from ethylbenzene (EB) is:



With P @ $0,42/kg, PO @ $1,41/kg, H_2O valueless, and O_2 @ $0,04/kg, the preliminary economics are:

$$(104)(0,57) + (58)(1.41) - (106)(0.48) - (32)(0,04) - (42)(0.42)$$

$$= \$141 - 70 = \$71/\text{kg mol}$$

Clearly this process appears to be more attractive economically, due to the value of propylene oxide.

The current styrene price, $0,57/kg, is at the low end of the historical range, which has been $ 0,50 to 1,50/kg. This is undoubtedly due to excess capacity, suggesting this is not a good time to invest in styrene capacity. The propylene oxide route looks much more attractive, but should be considered only after market studies for styrene and propylene oxide.

PROBLEM 4-4

We select the ethylbenzene dehydrogenation path, because it is most widely used. For a similar analysis of the propylene oxide path, see problem 4-14.

The chemistry of the process is considered more thoroughly. Reaction (1) is an equilibrium reaction; it does not go to completion. The reaction is conducted in the gas phase over a solid, ferric-oxide based, catalyst. It is carried out at about 1-atm total pressure (low pressure favors the products of this equilibrium) and a temperature around 600°C (1112°F). Steam is added to lower the partial pressure of the products, thereby further favoring their formation. The reaction is endothermic, with the heat required being supplied by the steam. A steam-to-ethylbenzene molar ratio of 8 to 1 is representative. At typical process temperatures other reactions occur, including

$$\underset{}{C_6H_5{-}C_2H_5} \rightleftharpoons \underset{\text{benzene}}{C_6H_6} + \underset{\text{ethylene}}{C_2H_4} \qquad (2)$$

and

$$C_6H_5{-}C_2H_5 + H_2 \rightleftharpoons \underset{\text{toluene}}{C_6H_5{-}CH_3} + \underset{\text{methane}}{CH_4} \qquad (3)$$

Reactions (2) and (3) consume ethylbenzene without producing the desired styrene; they also produce by-products that must be separated from the product and disposed of or utilized.

The process feeds are ethylbenzene and steam and the products are condensed steam, styrene, benzene, toluene, hydrogen, methane, and ethylene. Including the steam, valued at $0.0[illegible]/kg, adds $0.[illegible] to the total feed cost, raising it to $50.7 per 104 kg of styrene product. The value of the products still exceeds that of the inputs, but barely. It is necessary to have information on the extent of each of the three reactions at the reactor conditions. These extents could be calculated by using complete reaction-kinetic information, and this probably would be done in a more complete design of the process.

Here the calculations are illustrated assuming a constant extent for each reaction. The values used for mass balance calculations are,

Fractional extent, reaction (1) = 0.47
reaction (2) = 0.025
reaction (3) = 0.005

Because it is a large volume product, run continously.

PROBLEM 4-4 (continued - 1)

The fractional extent-of-reaction is the fraction of **the key reactant**, ethylbenzene, in the reactor feed that reacts while passing through the reactor. In this example, 0.5 mol of ethylbenzene are consumed per 0.47 mol of styrene produced.

A total fractional extent-of-reaction less than 1.0 (0.5 in this example) means that not all of the reactant in the feed reacts, so unreacted reactant is present in the reactor-product stream. The latter stream contains ethylbenzene, all the reaction products, by-products, and steam. It is essential to separate the main product, styrene, from the rest of this mixture. Economically it is desirable and generally it is necessary to recover unreacted reactants and recycle them to the reactor. By-products must be separated and, if possible, effectively utilized. The **condensed** steam must be separated from the product stream and removed.

Mass balances for the process are recalculated using the extent-of-reaction values from above and assuming that all unreacted ethylbenzene (EB) is recycled and converted to products. On the basis of 1 kg-mol (104 kg) of styrene product, the calculations are:

$$\begin{aligned}\text{Feed EB flowrate} &= (1\ \text{kg-mol styrene/h})\\ &\quad *(0.5\ \text{mol EB}/0.47\ \text{mol styrene})\\ &\quad *(106\ \text{kg EB/kg mol})\\ &= 112.8\ \text{kg/h}\end{aligned}$$

The EB feed rate to the reactor must be twice this, because only one-half of the feed reacts per pass and the EB recycle rate equals the feed rate. Thus,

$$\text{Reactor EB feed rate} = 225.6\ \text{kg/h}$$

$$\begin{aligned}\text{The steam feed rate is then} &= (225.6/106\ \text{mol EB/h})\\ &\quad *(8\ \text{mol steam/mol EB})\\ &\quad *(18\ \text{kg steam/mol steam})\\ &= 306\ \text{kg/h}\end{aligned}$$

The by-product benzene plus toluene and the fuel gas rates are calculated, from Eqs. (2) and (3), to be 5.2 and 3.6 kg/h, respectively. The raw materials and product values can be compared again. The benzene-toluene mixture is valued at \$0.2 /kg and the fuel gas at \$0.[illegible]5/kg (both based on heating value); condensed water has no value. The value of the feeds, ethylbenzene and steam, is \$57.3/104 kg of styrene. The value of the product plus the by-products and fuel gas is \$60.[illegible]/104 kg of styrene. The value of the outputs still exceeds that of the inputs, but the margin has narrowed.

The process, as identified **so far, is shown** in Fig. 4-4.

PROBLEM 4-4 (Continued-2)

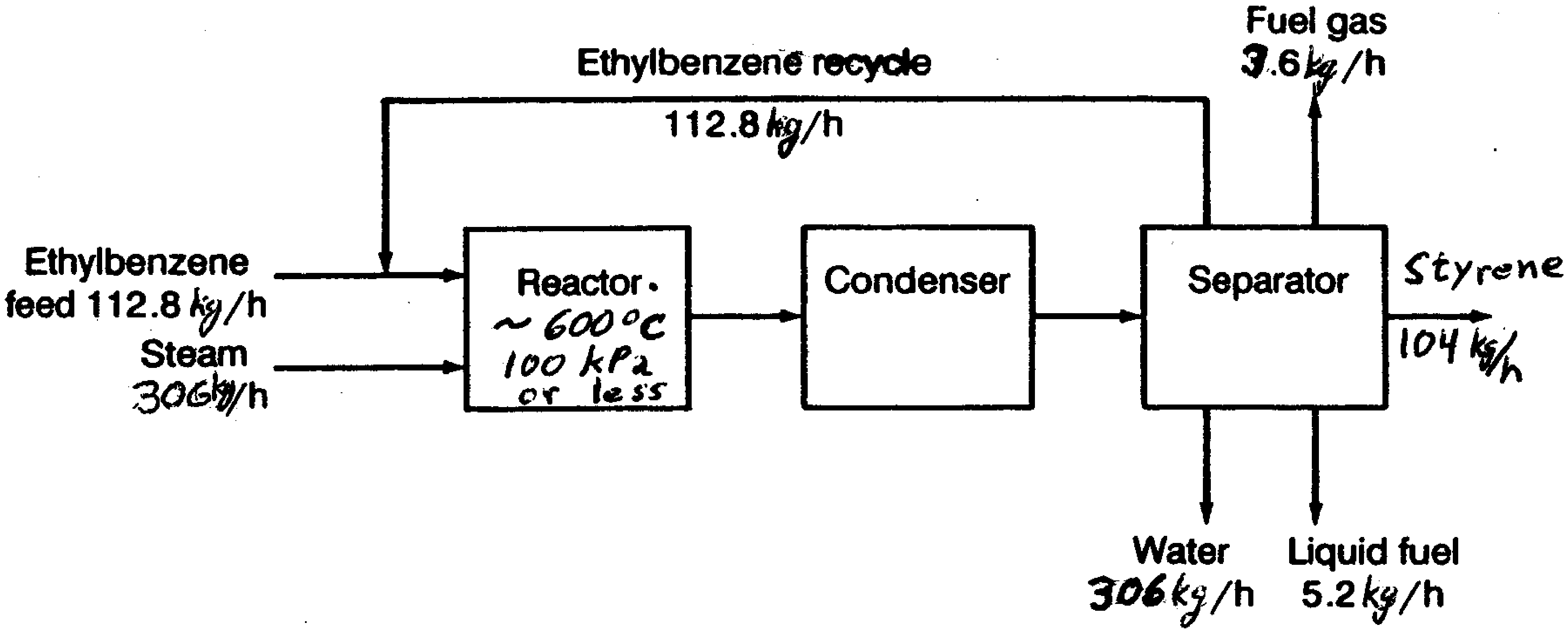


FIGURE 4-4 FUNCTIONS DIAGRAM

PROBLEM 4-5

One reactor is needed. The reaction is endothermic. The steam serves both as a diluent and a heat source. A packed bed of catalyst will serve as the reactor. If the reaction is adiabatic, temperature will decrease along the bed. An alternative is to have several beds in series with reheating between the beds.

Decision: Use adiabatic fixed bed reactor.

Separation processes. The **separation** after condensation into a water **phase, a** hydrocarbon phase, and a gas phase is accomplished by gravity in one or two holding tanks (other phase-separation methods are available and could be considered in a more detailed design). The liquid hydrocarbon phase needs to be separated into three fractions: (1) benzene plus toluene, (2) ethylbenzene, and (3) styrene. Experience shows that components with boiling points near ambient temperature (between, say, 0 **and** 200°C) and with adequate differences between their **boiling** points (at **least 5°C) are** usually most economically separated by **distillation.**

Normal boiling point values (°C) for the components of this system are:

hydrogen	−252.5	water	100
methane	−161.5	toluene	110.6
ethylene	−104	ethylbenzene	136
benzene	80.1	styrene	145

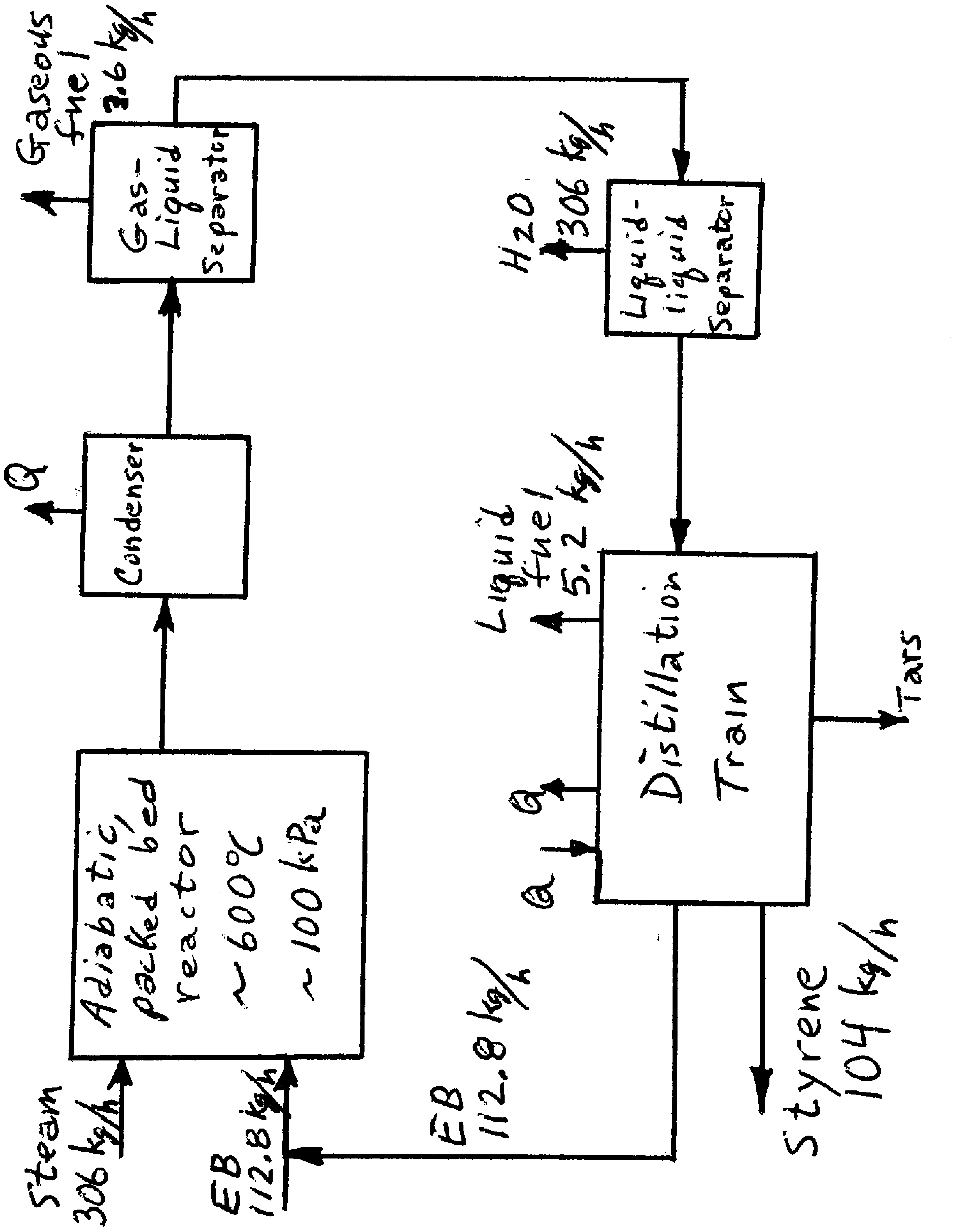
These values show that, with ambient cooling at 1 atm, hydrogen, methane, and ethylene are difficult to condense, but **that steam,** benzene, toluene, ethylbenzene, **and** styrene are easily condensed. **Condensation** will separate the latter **five** components as liquids and leave **the first three** in the gaseous state. Gaseous mixtures are very difficult to separate; when they are combustible, as here, experience indicates that it is usually best to use them as a fuel. Property data also show that water and the liquid hydrocarbons are highly insoluble **in each** other and thus will split into two liquid phases **that** can be separated **easily.**

Decision: Condense benzene and higher boiling components. Use holding tank to separate non-condensibles from liquids.

PROBLEM 4-5 (Continued-1)

<u>Decision:</u> Separate water from hydrocarbon liquids via holding tank (decanter).

<u>Decision:</u> Separate the hydrocarbon liquids by a distillation train.



OPERATIONS DIAGRAM

PROBLEM 4-6

An ordinary fractional distillation column yields two products; two distillations are required to produce the three hydrocarbon products in this process. Experience again provides guidance in planning this separation: First, remove the lower boiling component (benzene and toluene) and, last, make the most difficult separation (closest boiling points—ethylbenzene and styrene). Another lesson of experience is to remove a desired product (styrene) finally as a distillate (lower boiling) product. That is not possible with this mixture as described; however, experience also shows that organic reactions almost always generate higher boiling "tars" that need to be removed. Thus, a final distillation step, with styrene as the distillate product and tar (not included in the material balances) as the bottom product, is recommended. Another complicating factor here is that styrene polymerizes when heated. In order to avoid significant polymerization, styrene distillation temperatures are lowered by operating under vacuum.

Decision: Use 3 distillation columns.

The first column can be at or near atmospheric pressure, as the dilution of the styrene plus added polymerization inhibitors will limit styrene polymerization. The second column, EB-styrene separation needs to be under vacuum. For example at 13.3 kPa (100 mm Hg) absolute pressure, styrene boils at 82°C. This will be the reboiler temperature. The third column also needs to be under vacuum; at 5.3 kPa styrene boils at 60°C.

Packed distillation columns should be used to reduce pressure drops and residence times. A vacuum system, e.g. steam-jet ejectors, is needed.

Separation processes in reality do not achieve perfect separation as we have assumed in the mass balances. Actual separation-product stream compositions must be specified if the separations are to be designed in more detail. Here again experience, as well as product and by-product specifications, would be used to establish these specifications. Equipment design is not included in this example, so these specifications are not discussed.

PROBLEM 4-6 (continued-1)

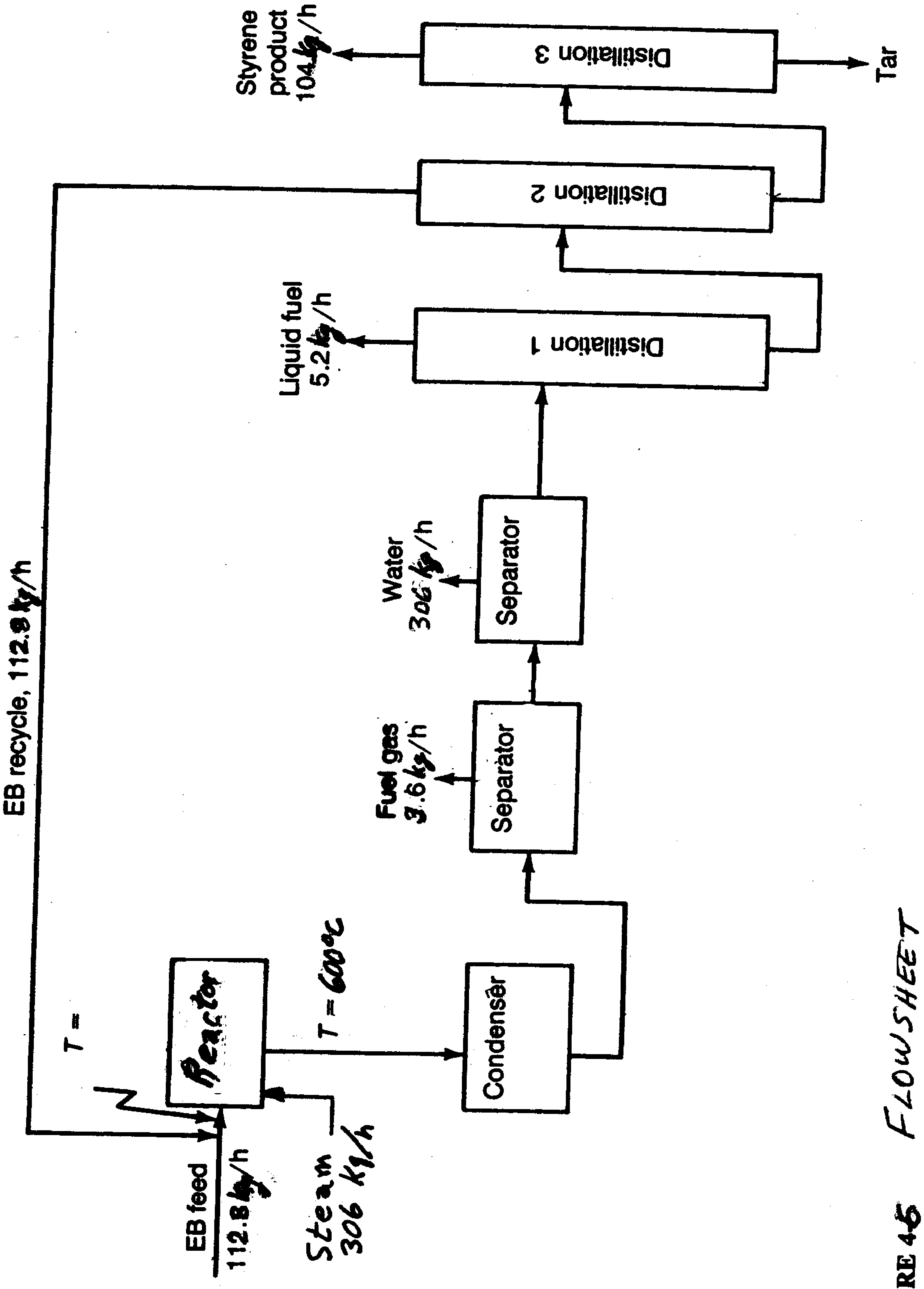


FIGURE 4.6 FLOWSHEET

PROBLEM 4-6 (continued - 2)

Heat integration. Both heating and cooling are needed in this process. Since supplying and removing heat is expensive, it is desirable to heat and cool using heat exchange between process streams. The purpose of heat integration is to satisfy the process heating and cooling requirements as economically as possible.

Required temperatures are specified; for example, a reactor outlet temperature of 600°C (1112°F) is needed. The temperature and energy requirements for the heating and cooling loads are matched as closely as possible. Styrene reactors typically operate adiabatically (no heat is added), causing the temperature to drop as the endothermic reaction proceeds. The reactor-inlet temperature required to achieve the specified outlet temperature is calculated from a reactor energy balance,

$$\text{Enthalpy of products} = \text{enthalpy of reactants} - \text{heat of reaction}$$

Enthalpies of reactants and products are expressed with respect to a reference temperature of 25°C (77°F), and the heat of reaction is available at 25°C:

$$[m * C * (600 - 25)]_p = [m * C * (T - 25)]_r - m * \Delta H_R$$

where m is the mass flow rate (kg/h), C is the heat capacity (kJ/kg-°C), $m * \Delta H_R$ is the total heat of reaction (kJ/h), and subscripts p and r denote products and reactants, respectively. The reactant and product mass flow rates are both 531.6 kg/h including steam from the earlier mass balances. Using 2.2 kJ/kg-°C for the heat capacity of both products and reactants, 117,500 kJ/h as the total heat of reaction (styrene molar flow rate * heat of reaction for styrene), and solving this equation for T gives a required inlet temperature of 700°C for the reactants.

PROBLEM 4-6 (continued - 3)

The ethylbenzene feed needs to be heated and the reactor effluent cooled. Heat exchange between them is feasible and desirable. Setting an approach temperature for a countercurrent exchanger (temperature difference between heated feed and reactor effluent) at 30°C, fixes the heated feed temperature at 570°C (1058°F). An energy balance for the heated feed and steam mixture establishes the required steam temperature:

$$[m * C * (T - 700)]_s + [m * C * (570 - 700)]_{EB} = 0$$

where the subscripts s and EB signify steam and ethylbenzene, respectively. Using $C_s = 2.5$, $C_{EB} = 2.5\ kJ/kg\text{-}°C$, and the flow rates calculated above gives

$$T(\text{of superheated steam}) = 796°C$$

An energy balance around the feed-effluent heat exchanger,

$$[m * C * (600 - T)]_{\text{effluent}} = [m * C * (570 - 60)]_{\text{feed}}$$

where $m*C_{feed} = (225.6)(2.5)$ and $m*C_{effl.} = (2.2)(531.6)$

$T = 354°C$ as the outlet temperature of the effluent

The reactor effluent must reach about 40°C (104°F) for condensation. Thus, further cooling is required, by cooling water or other process loads. Heating ~~and cooling~~ for the distillation columns represent such possible loads.

Figure 4-6, on the next page, shows the heat integrated flowsheet.

The analysis evokes the question of utilizing the heat in the 354°C effluent. The only places where heat is needed are the distillation column reboilers. Analysis of the distillations to find reboiler heat loads is needed, but is not attempted here. Such application would result in some change in the flowsheet.

The earlier decisions should be revisited. Changes these would change the flowsheet.

PROBLEM 4-6 (continued - 4)

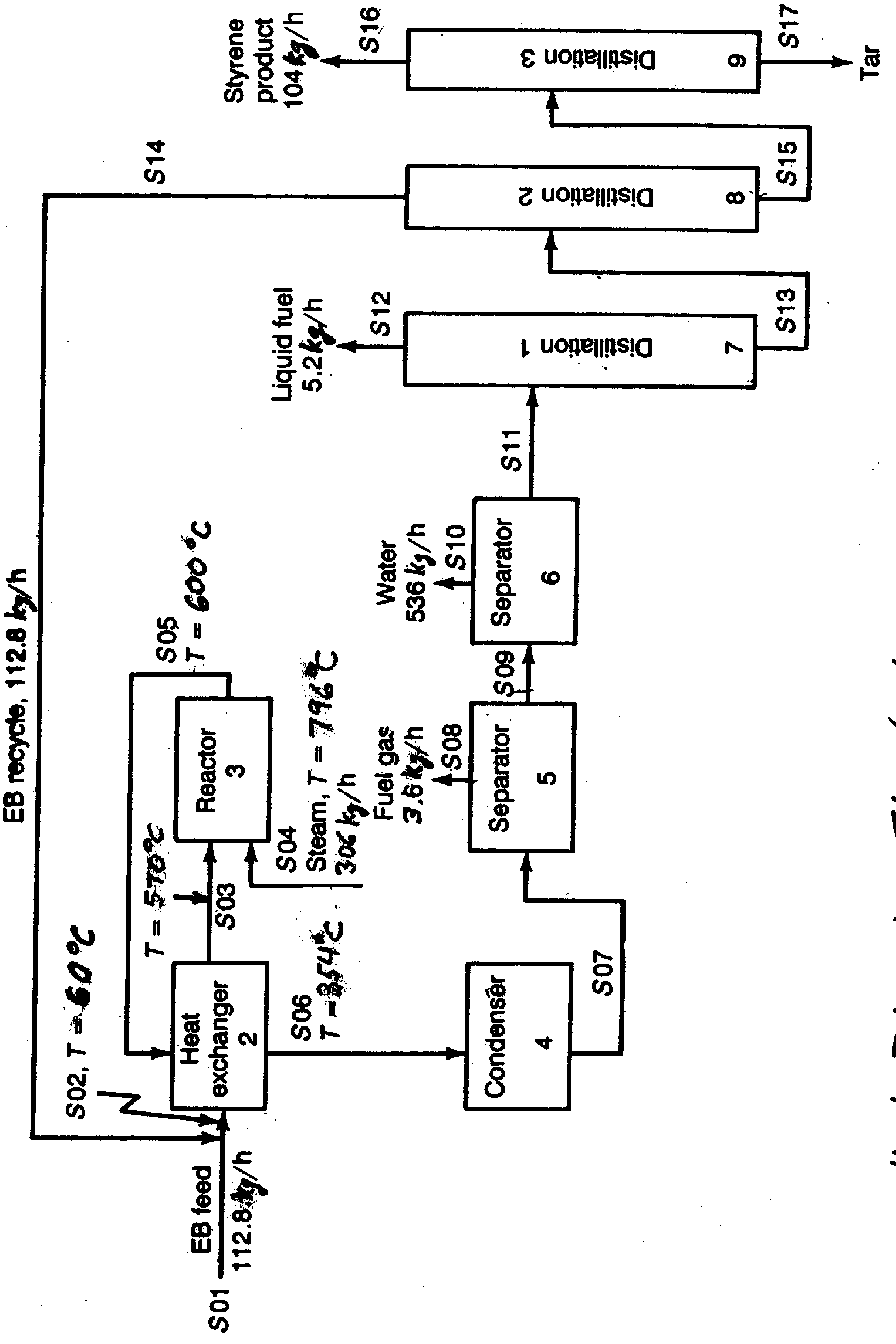


FIGURE 4. Heat Integrated Flowsheet

PROBLEM 4-7

Based on *Kirk-Othmer Encylcopedia*, assume that a reasonable catalyst lifetime is 18 months. Assume that in order to achieve a continuous operation, twice the amount of catalyst needed in the reactor is purchased.

Basis: 104 kg/h of styrene product. This requires 225.6 kg/h of reactor feed (see problem 4-6).

Given: 1 kg of catalyst is required in the reactor per 100 kg/h of reactor feed.

Thus,

(1/100)(225.6 kg /h of reactor feed) = 2.256 kg catalyst/104 kg styrene

Assume that twice the required amount of catalyst is purchased, so that there is sufficient catalyst to achieve continuous flow of product (e.g. by using 2 reactors).

The purchase cost of catalyst = (4.512 kg cat.)(5\$/kg cat.)/(104 kg styrene/h)

= \$0.217/(kg styrene/h)

Based on the assumed 18-month catalyst lifetime, the operating lifetime of the catalyst is

(18/12)(365)(24) = 13,140 h

Purchase cost of cat. = (\$0.217)/(kg styrene/h)(13,140 h) = \$1.65*10^{-5}/kg styrene

The regeneration cost = (2.256 kg cat.)(\$0.15/kg)/((104 kg styrene/h)(800h))
= \$4.1*10^{-6}/kg styrene

Comparing these costs with the styrene value of \$0.57/kg and ethylbenzene cost of \$0.48/kg it is clear that, for the stated and assumed conditions, the catalyst cost does not have a significant impact on the economics of the process.

PROBLEM 4-8

A simulation can be performed using one of the overall process simulators listed in Appendix F, including Aspen Plus, CHEMCAD, Hysis, or others.

In order to perform the simulation, the components and the physical property routine must be specified. The main components in the styrene process are ethylbenzene, benzene, ethylene, toluene, methane, styrene, hydrogen, and H_2O. For example, the Peng-Robinson method for properties could be used.

The operations needed include: mixing, a reactor, splitting, heat exchange, cooling and condensing, distillation, recycle, and steam jet ejectors. The ethylebenzene conversion per pass is specified, the styrene selectivity can be estimated along with the extent of the benzene and toluene reactions. These could be calculated with a kinetic based reactor model, and should be for more detailed design. A simplified distillation model could be used, such as one based on the FUG method. Pressures must be established for the distillation columns. Split and recovery estimates are needed for the components.

The simulator will calculate material and energy balances. These would be useful in looking for heat integration opportunities. Approximate equipment specifications would also be established.

PROBLEM 4-9

See problem 4-3 for the two significant processes for styrene, and for pertinent sources.

Ethylbenzene is the only raw material for the dominant dehydrogenation process discussed in problems 4-3 through 4-8. Ethylbenzene and propylene are the raw materials for the route of secondary importance, the propylene oxide process.

Ethylbenzene is made from benzene and ethylene by

$$C_6H_6 + C_2H_4 \longrightarrow C_6H_5\text{-}C_2H_5$$

See Chapter 3 for a detailed description of the ethylene-production process.

Preliminary economics, on the basis of one kgmol of ethylbenzene show:

Product value = ($0.48/kg)(106 kg) = $50.9
Reactant values, Benzene value = ($0.40/kg)(78) = $31.2
Ethylene value = ($0.53/kg)(28) = $14.8
Product value – Reactant values = $ 4.9

This result illustrates that, if the producer must purchase the benzene and ethylene at these prices, the operation is only marginally promising.

For the propylene oxide process, the economics of ethylbenzene production are the same as above. Propylene is a co-product of the ethylene process described in detail in chapter 3, and is economically favorable as shown there. The propylene oxide process itself is economically promising as shown in Problem 4-3.

In fact, the standard in the industry is that styrene production is integrated from benzene and the ethylene raw material (e.g. propane) through styrene (and propylene oxide for the second process). Styrene production is part of an integrated petrochemical process.

PROBLEM 4-10

Based on problem 4-9, there is no strong incentive to integrate ethylbenzene production with styrene production, for either the dehydrogenation or the propylene oxide route. However, the reality of the situation is that they are always integrated. Styrene is by far the major market for ethylbenzene, and the economics are favorable when the styrene producer controls raw material supply. The price of ethylbenzene on the open market does not reflect its value to the integrated producer.

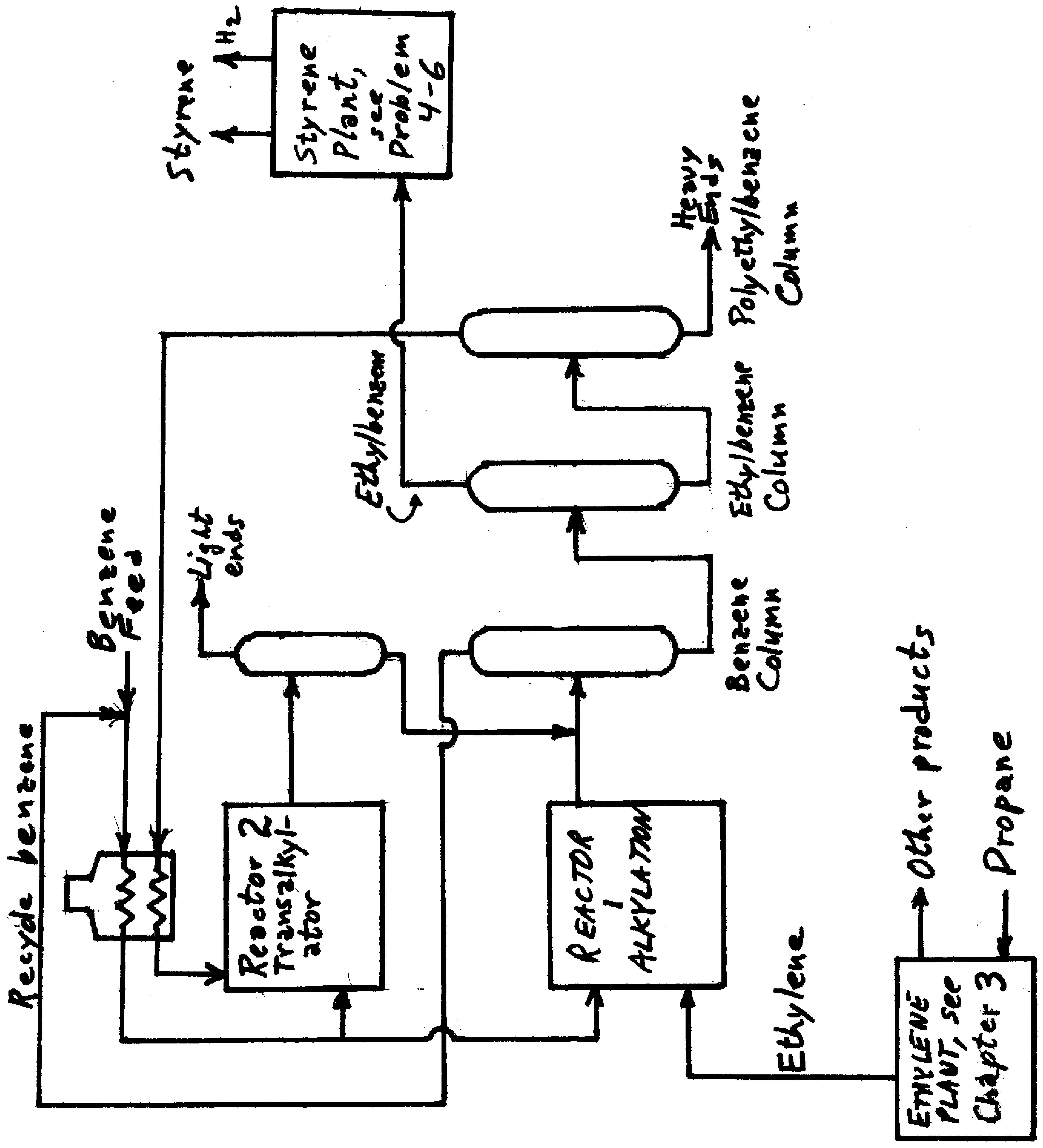
Ethylene and propylene are co-products of ethylene production (see Chapter 3), and this is an economically favorable process. So, it follows that the ethylene (and propylene), benzene, and ethylbenzene production would be integrated with the styrene production.

See problem 4-12 for preliminary economics.

PROBLEM 4-11

Reaction 1. Vapor-phase alkylation of Benzene + ethylene to ethylbenzene over zeolite catalyst.

Reaction 2. Vapor-phase transalkylation of benzene with polyethylbenzenes (by-products of 1.) over zeolite.



4'.21

PROBLEM 4-12

The preliminary economics of the overall process shown in Problem 4-11 are shown.

$$C_3H_8 + C_6H_6 \longrightarrow C_6H_5\text{-}C_2H_3 + H_2 + CH_4$$

Product value (neglect H_2 and CH_4) = (\$0.57/kg)(104 kg) = \$59.3

Reactant values: Propane = (\$0.27/kg)(44 kg) = \$11.9
Benzene = (\$0.40/kg)(78kg) = \$31.2

Value of products – reactants = \$16.2

Note: Benzene price is from www.icislor.com, a good source for chemical and petrochemical prices.

This result shows a significant economic improvement compared to the result in problem 4-9. Similarly, the propylene oxide pathway, shown in problem 4-3 to be favorable, also shows improvement.

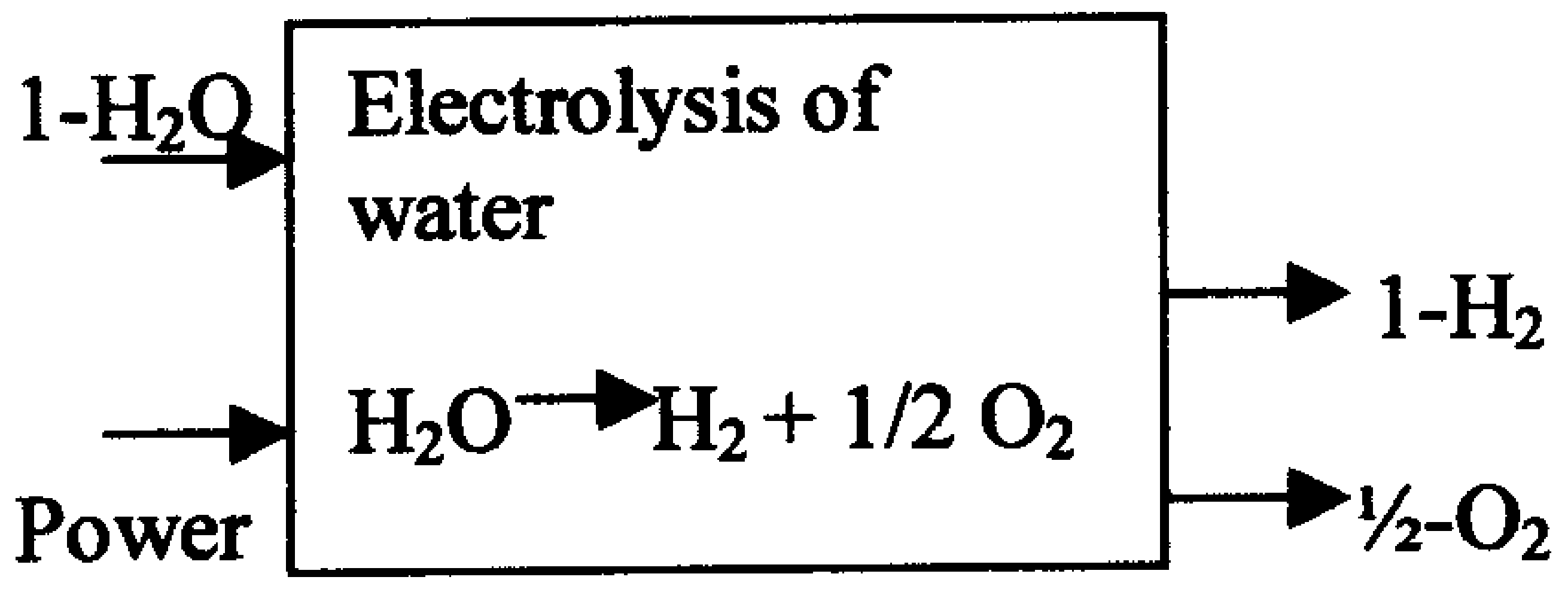
4.22

PROBLEM 4-13

Sources: *Kirk-Othmer Encyclopedia of Chemical Technology*, 4th ed., and *Ullman's Encyclopedia of Industrial Chemistry,* 5th ed.

For either of the rates suggested, a continuous process should be used. The potential of hydrogen as a fuel for fuel cells would require very large production rates and large plants. An alternative that has been suggested is small electrolysis plants located at refueling stations.

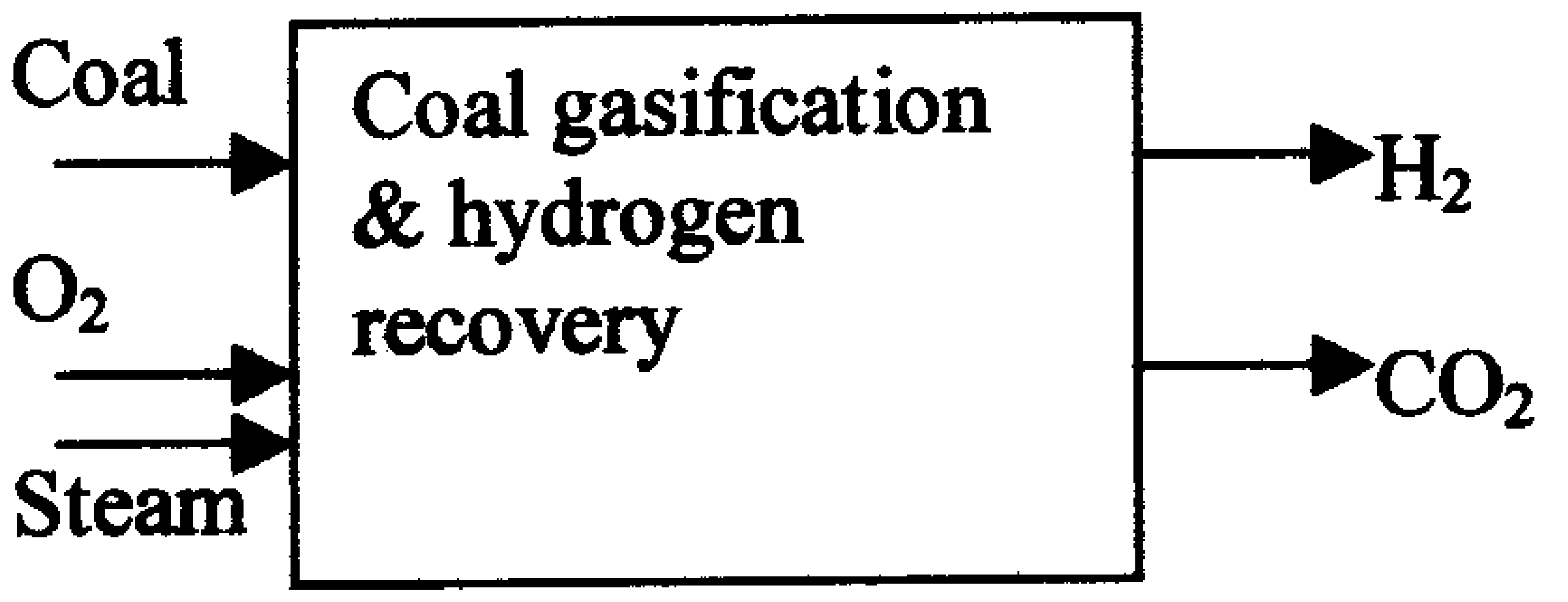
Water. Hydrogen can be obtained from water by any decomposition process, but the most likely candidate is electrolysis. An I/O diagram and preliminary economics are shown.



	Price	Amount/ kgmol H_2	Cost/ kgmol H_2
H_2O	negl.	18 kg	negl.
Elec.	\$0.05/kWh	50 kWh	\$-2.50
H_2	\$0.67/kg	2 kg	\$ 1.34
O_2	\$ 0.04/kg	16 kg	\$ 0.64
		Net	\$-0.52

A price for hydrogen difficult to come by, but \$0.67 is at the high end of the range (\$0.01 to 0.06 per standard m^3) for pipeline hydrogen from Kirk-Othmer. This pathway is not economically attractive for pipeline-scale quantities.

Coal. Coal may be gasified to produce a gas containing CO, CO_2, and H_2. Hydrogen may be recovered from this gas.

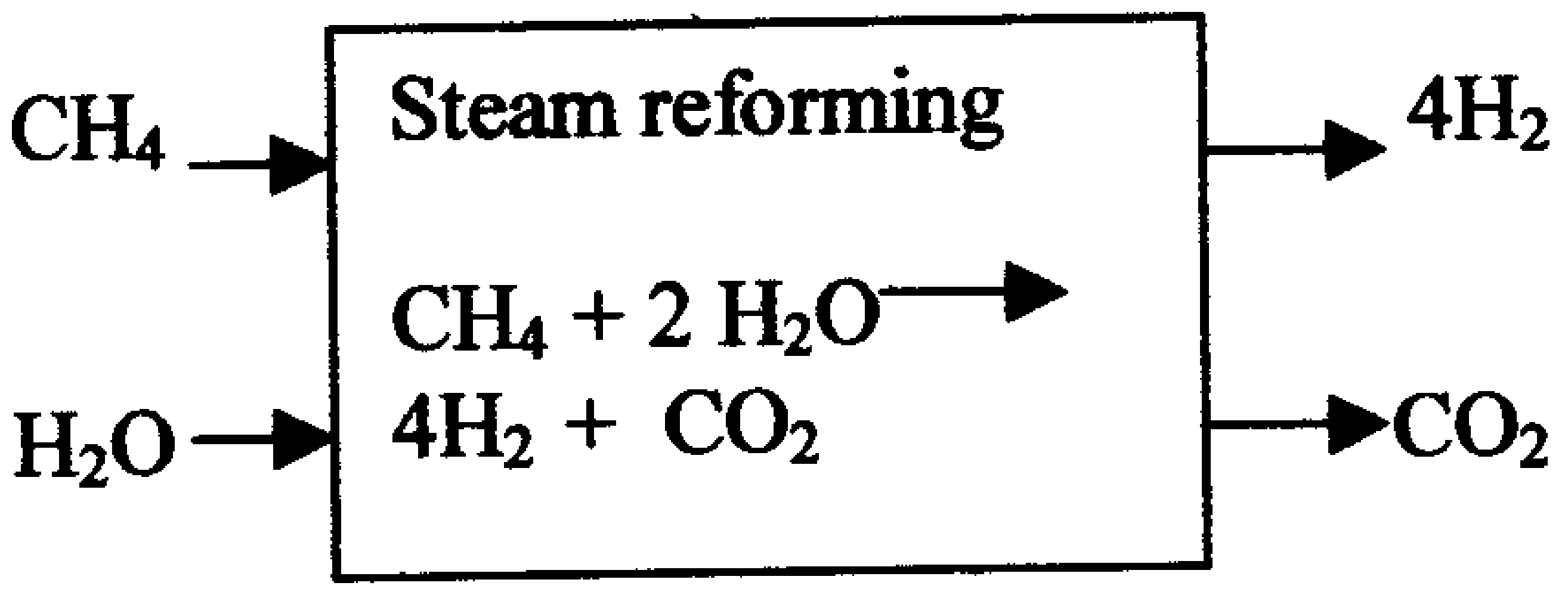


	Price	Amount/ kgmol/H_2	Cost/ kgmol H_2
Coal	\$0.055/kg	~17	\$-0.94
O_2	\$0.04/kg	~ 4	\$-0.16
Steam	\$0.008/kg	~20	\$-0.16
H_2	\$0.67/kg	2	\$ 1.34
		Net	\$ 0.08

This pathway is marginally economically attractive with no processing cost included.

PROBLEM 4-13 (continued-1)

Natural gas. Natural gas (NG, mainly methane) is converted to a gas containing primarily CO_2 and H_2 by reacting with steam, the steam-reforming process.



	Price	Amount/ kgmol H_2	Cost/ kgmol H_2
NG	\$0.13/kg	4 kg	\$-0.50
Steam	\$0.008/kg	9 kg	\$-0.07
H_2	\$0.67/kg	2 kg	\$ 1.34
		Net	\$0.77

This pathway looks potentially economical. Indeed, this well developed process is the way in which industrial hydrogen is produced.

PROBLEM 4-14

Note corrections: The first four words should be replaced by "For the second…" The sentence beginning, "100 percent separation of all ethylbenzene…" should read, "100 percent separation and recycle of all ethylbenzene from the reactor exit stream." The stated catalyst life of 1100 hrs is actually the period between regenerations. Assume a catalyst life of 1 year. Due to lack of thermodynamic data for several components, heat integration should not be required.

See problem 4-3 for I/O diagram and preliminary economics.

First consider the catalyst cost for the dehydration of α-methylbenzyl alcohol. Given that 1 kg of catalyst is required per 100 kg/hr of feed, the amount of catalyst required per 100 kg/h of styrene product is approximately (122/104). Three catalysts are required, thus,

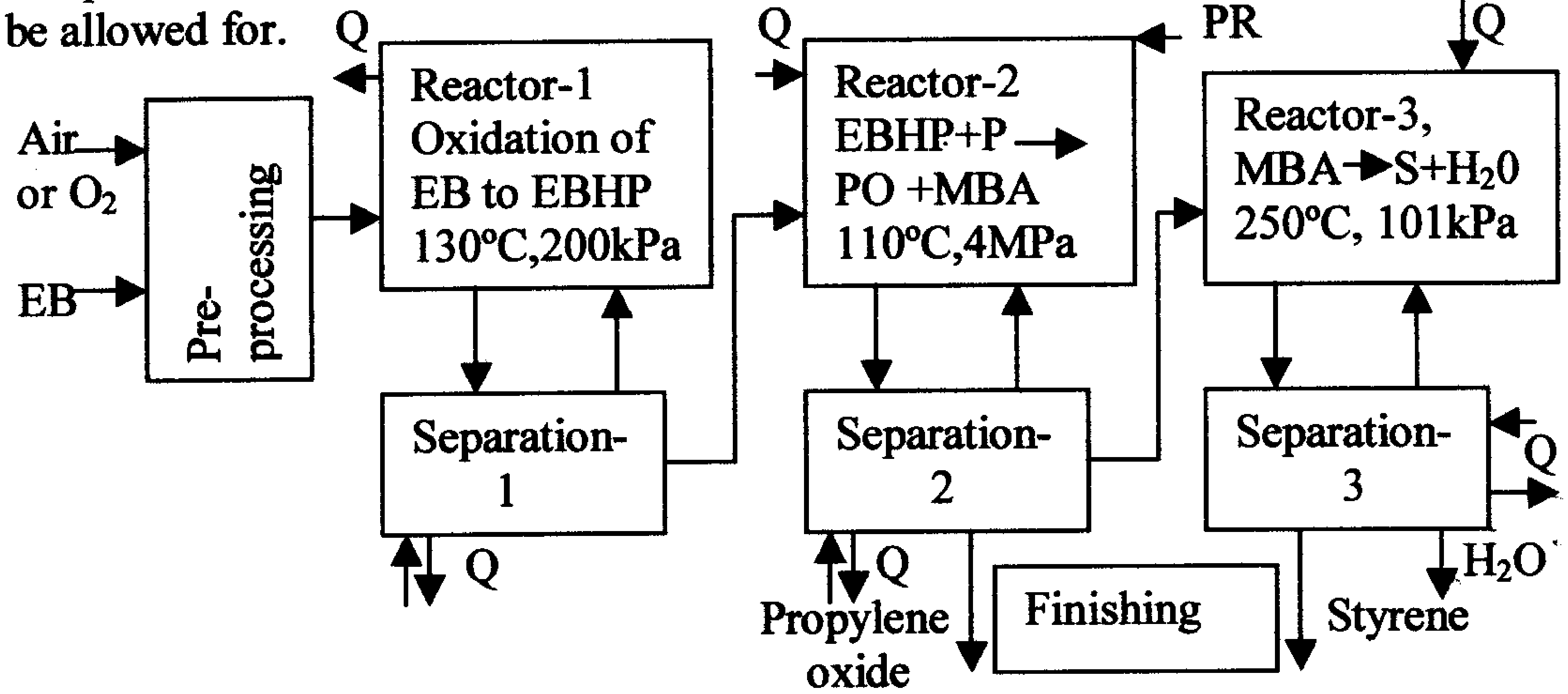
Catalyst required = (3)(122/104 kg feed/kg product)(1kg cat./100kg feed/h)
= 0.0352 kg catalyst/kg styrene product/h

At 1100 h per regeneration, there would be 7 regenerations over the assumed one year lifetime of the catalyst. Further assuming that twice the total amount required catalyst is purchased, the catalyst cost becomes

Catalyst cost = (2)(0.0352 kg cat./kg styrene/h)($6.6/kg cat.)(1/8760 h) +
(7)(0.0352 kg cat./kg styrene/h)($0.25/kg cat.)(1/8760 h)
= $$6*10^{-5}$/kg styrene

Comparison with problem 4-3 shows that this cost does not significantly affect the process economics.

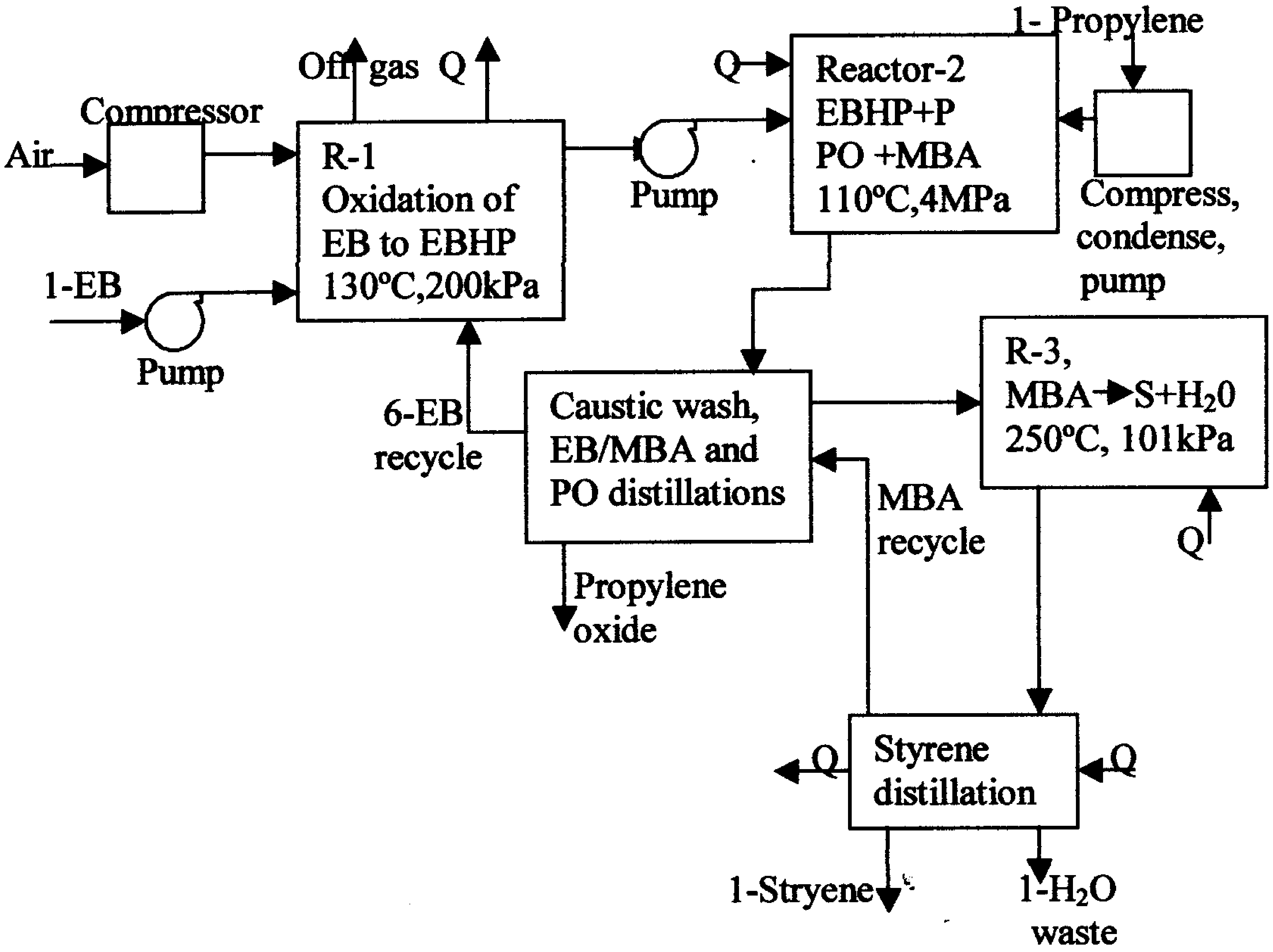
a) Functions. Three separate reactors are required for this process. Separation and purification is required after each reactor. Preprocessing and finishing should be allowed for.



PRpropylene, PO=propylene oxide, EB=ethylbenzene, EBHP= ethylbenzene hydroperoxide, and MBA=α-methylbenzyl alcohol, S=styrene

HOMEWORK 4-14 (continued-1)

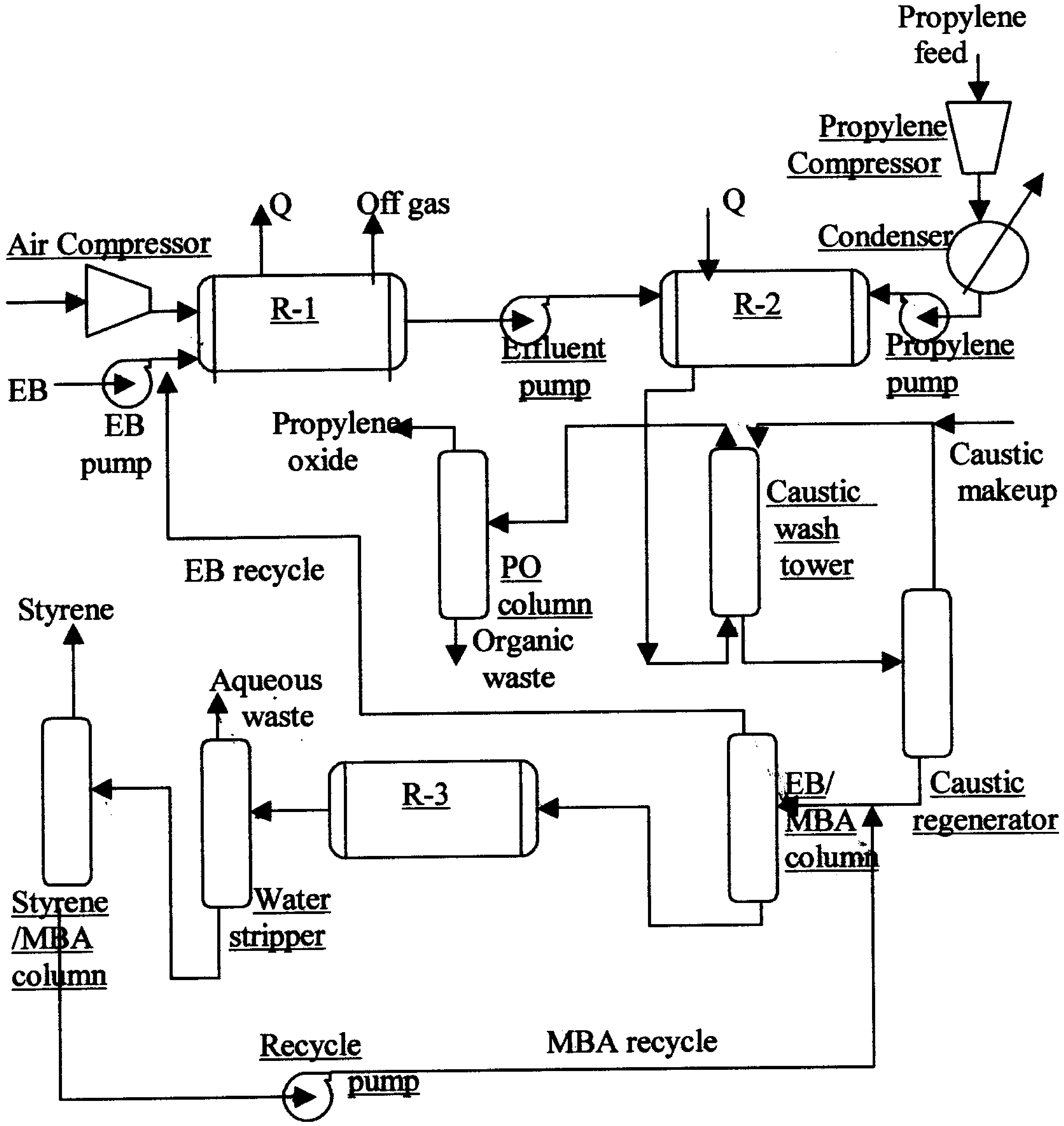
b) Operations. Air is used for the liquid-phase oxidation of EB, as the nitrogen moderates the temperature. Air needs to be compressed and EB pumped into reactor-1. EB conversion is only about 15 percent in reactor-1, to minimize byproducts. The products of reactor 1 are sent directly to reactor 2. Unreacted EB is recycled after reactor-2. Conversion of EBHP is essentially complete. Caustic solution is used to absorb all of the organics except PO, which is purified by distillation. EB and MBA are separated by distillation, the EB recycled and the MBA goes to R-3. From R-3, styrene is recovered and purified by distilation and MBA is recycled to the first separation process. Insufficient data available in the literature and on the web to obtain the heat duty values, Q.



c. Flowsheet. The operations diagram is coverted to a flowsheet by defining the recovery and separation steps. See the next page. Because of the lack of heat of reaction data, heat integration has not been accomplished in this study.

PROBLEM 4-14 (continued-2)

d. The economics has been examined in problem 4-3 and above.



PROBLEM 4-15

Sources: www.chemicalmarketreporter.com, *Kirk-Othmer Encyclopedia of Chemical Technology.*

a) The major process for acrylic acid production is from propylene and oxygen:

Reaction 1, $\underset{\text{propylene}}{CH_2{=}CHCH_3} + O_2(\text{air}) \rightarrow \underset{\text{acrolein}}{CH_2{=}CHCHO} + H_2O$ Exothermic

Reaction 2, $CH_2{=}CHCHO + \frac{1}{2}\,O_2\ (\text{air}) \rightarrow \underset{\text{acrylic acid}}{CH_2CHCOOH}$ Exothermic

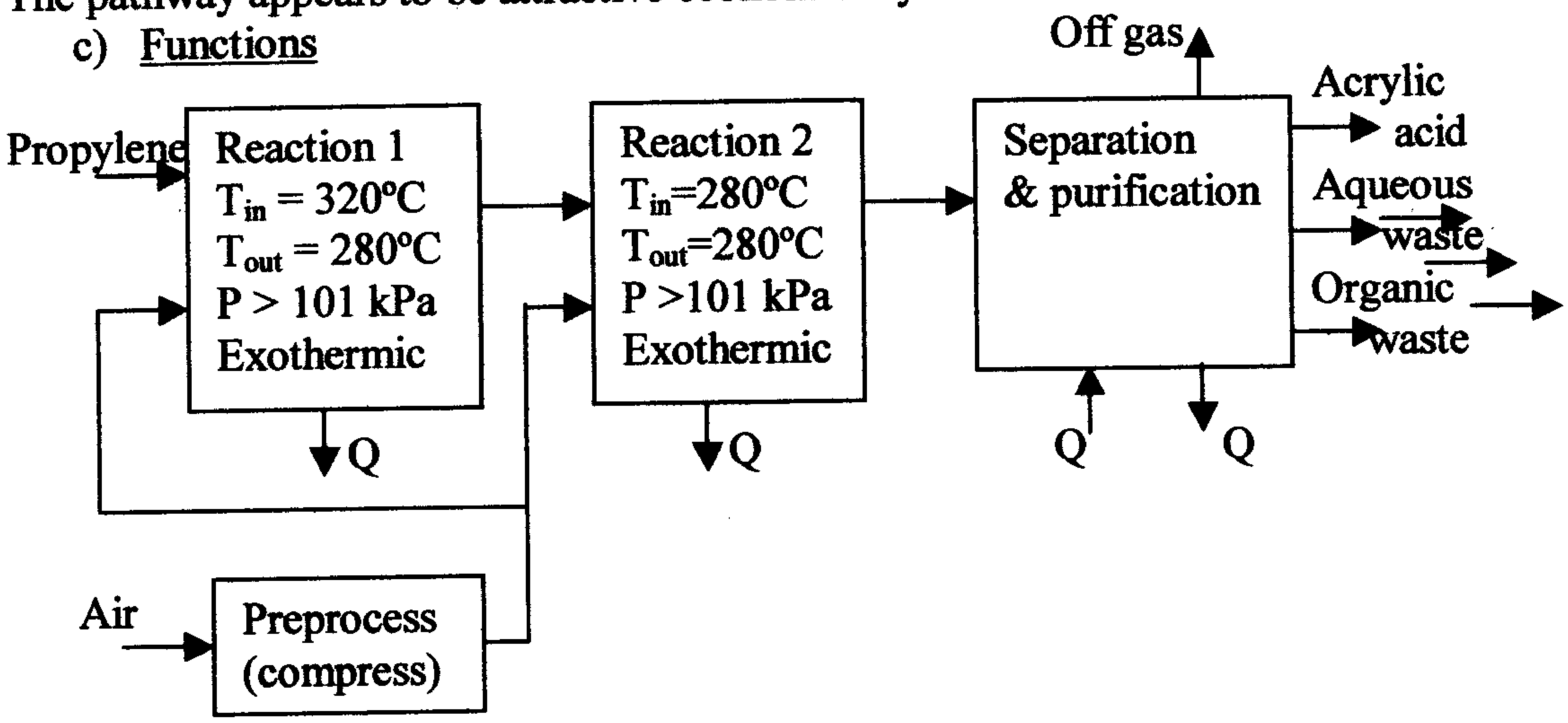
Acrylic acid is a high volume chemical, continuous processing is used. Similar catalysts are used for both reactions, but the catalysts and conditions are sufficiently different that the reactions are conducted separately. Bismuth molybdate and molybdenum vanadium oxides typically are the bases for the catalysts in the first and second reactions, respectively. Effluent from the first reactor can go directly to the second reactor without processing. Fixed bed, shell-and-tube, solid-catalyst reactors are used for the gas-phase reaction. Reactors are cooled with circulating molten salts. Additional process information is presented as needed.

b) Basis: one kgmol of acetic acid

	Price	Amount	Value
Propylene	\$0.42/kg	42 kg	\$17.64
O_2 (air)	negligible	48 kg	negl.
Acetic Acid	\$1.0/kg	72 kg	\$72.00
H_2O	negligible	18 kg	negl.

The pathway appears to be attractive economically.

c) Functions



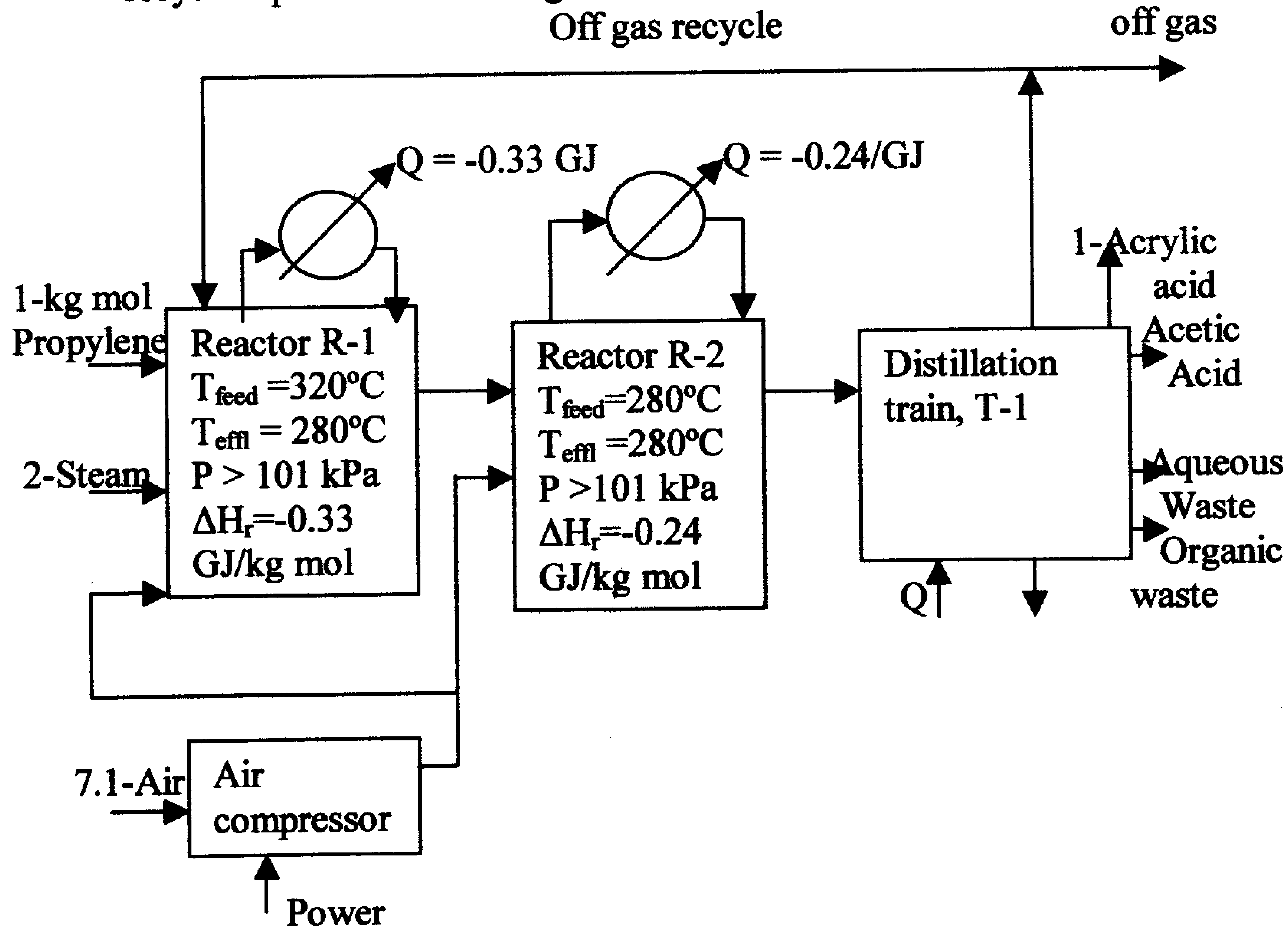
PROBLEM 4-15 (continued-1)

Operations. Steam is added to the reactor feed as a diluent. Molten salt is circulated through each reactor and cooled externally, usually by generating low-pressure steam. Conversion of ethylene and acrolein is nearly complete and 85 to 90 percent is converted to acrylic acid with the remainder being mostly acetic acid, a valuable byproduct. Depending on the conversions, recycle of off gases may or may not be necessary.

Decision: Design for 90% conversion to acrylic acid and the remainder acetic acid.

The separation of products is usually accomplished by first absorbing reactor-2 effluent into water, followed by extraction of the organics into a solvent such as butyl acetate, and distillation to recover the solvent, acrylic acid and acetic acid, and aqueous and organic wastes. However, if relatively high propylene and low steam concentrations in the feed are used, extraction can be dispensed with. In the interest of minimization of processing, this design will be employed, but it does require recycle.

Decision: Separate by distillation, do not use solvent extraction, but recycle a portion of the off gas.



PROBLEM 4-15 (continued-2)

Flowsheet. The distillation train must remove water, and recover and purify acetic and acrylic acids. Water and the organics form a heterogeneous azeotrope. The in order of increasing normal boiling points, in °C, is:

non-condensables, mainly nitrogen	
acrolein	52.5
water	100
acetic acid	118
acrylic acid	141

To prevent reactions, especially polymerization of the acrylic acid, the temperature in the distillation train should be kept below 100°C. At 100°C the vapor pressure of acrylic acid is 24 kPa.

Decision: Operate the distillation columns at about 16 kPa (90°C b.p. for acrylic acid). Remove water in the first column.

The overhead from the first column, C-1 contains the components more volatile than acetic acid. Following condensation, the two-phase liquid plus gas goes to a tank where sufficient holding time allows separation into a gas, an aqueous phase, and an organic phase. The gas is vented, part of it to be recycled and the rest sent to a flare. The organic phase is completely refluxed to the column. The aqueous phase is a waste stream, but it probably should be sent to a stripping column to reduce the organic content, and the organics sent back to column C-1. An aqueous stripping column is not shown in the flowsheet on the next page.

The remaining organic phase consists of acetic and acrylic acids, and less volatile components. Following the heuristic to remove largest quantities first, the second column, C-2, should separate into two streams, predominantly acetic and acrylic acids.

Decision: Next separate between acetic and acrylic acids.

Acetic acid is the distillate product, and should be obtainable in adequately pure form to meet specifications. The acrylic acid bottom product contains nearly all of the heavier components from the reactions. It needs to be distilled in order to meet specifications, as suggested by the heuristic to finally recover product streams as distillate products, and this is done in column C-3.

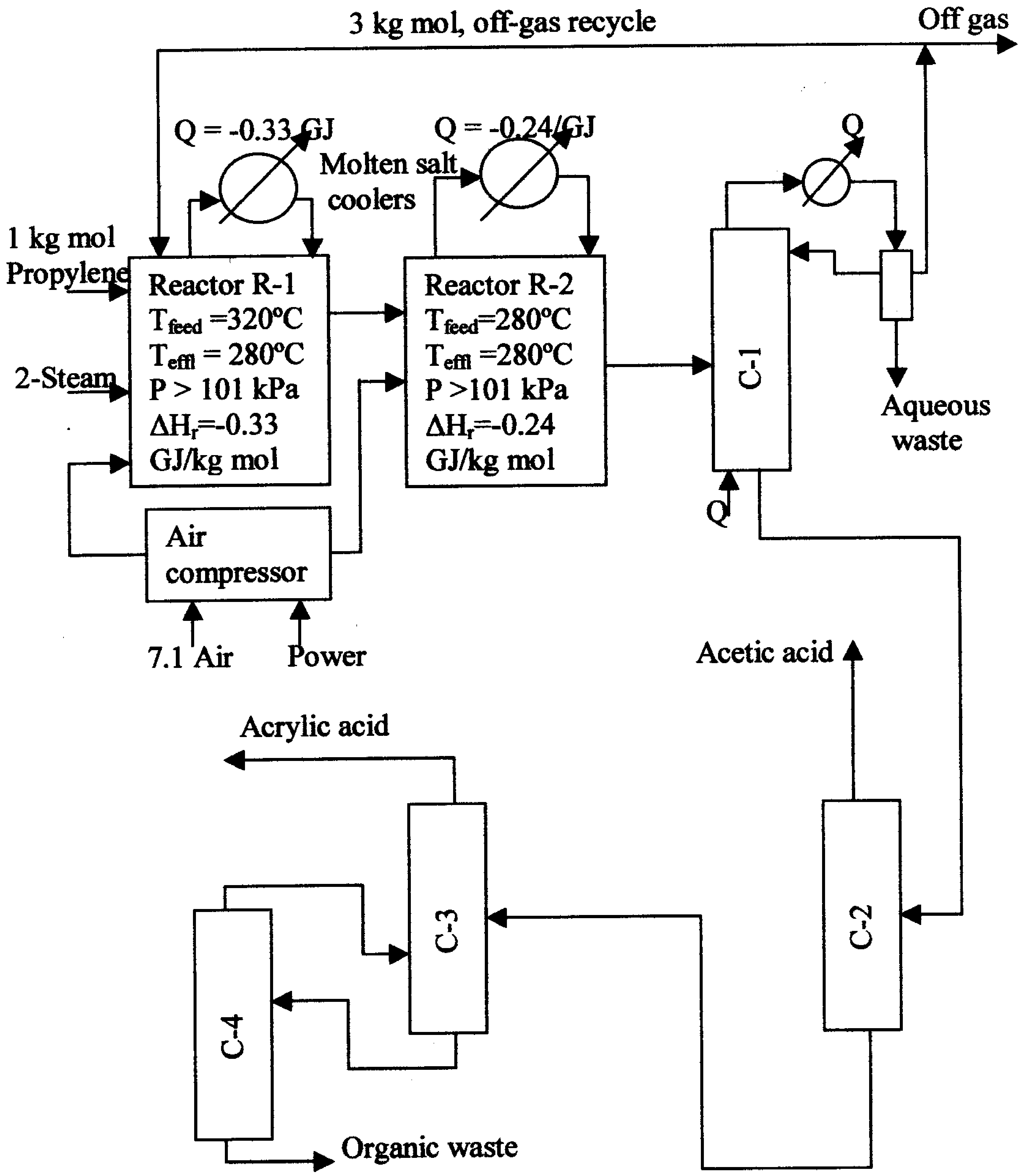
Decision: Distill the acrylic acid and recover it as a distillate product.

The bottom product from the acrylic acid column, C-3, is an organic liquid waste, but it is likely to still contain enough acrylic acid to make it worth while to send it to column C-4 to strip the acrylic acid. The acrylic acid is then recycled to the acrylic acid column.

Decision: Strip the C-3 bottom stream to recover acrylic acid values.

PROBLEM 4-15 (continued-3)

The flowsheet obtained is shown below.



PROBLEM 4-15 (continued-4)
d) The heat sources in the process are the two exothermic reactions. The column condensers are heat sources, but at temperature too low to be of much interest. The heat sinks are the distillation column reboilers. The temperatures and heat duties should be compared to see if some matching can be done, and surely some can. Generating steam in the coolers for the molten salts streams is one manner of heat integration, but it may be possible to use the molten salt streams directly as heat sources. Using the molten salt conserves some heat, because there is a temperature drop of the order of 25°C associated with every heat exchange. See Problem 4-6 for an illustration of heat integration with distillation.

e) A simulation can be performed using one of the overall process simulators listed in Appendix F, including Aspen Plus, CHEMCAD, Hysis and others.

In order to perform the simulation, components and the physical property package must be specified. The main components of this process are water, air, propylene, acrolein, acrylic acid, acetic acid, and acrylic acid dimer. The Peng-Robinson method could be used for example.

The operations needed include: two reactors, splitting, mixing, cooling, condensing, recycle and distillation. The propylene and acrolein conversions and product yields are to be specified. These could be calculated if sufficient kinetic data were available. A simplified distillation model, such as the FUG method, could be used. Split and recoveries need to be estimated.

The simulator will calculate material and energy balances. These would be useful in looking for heat integration opportunities.

PROBLEM 4-16

Sources: *Ullman's Encyclopedia of Industrial Chemistry*, 5th ed., *Chemical Market Reporter.*

a) The dominant chemical pathway for vinyl acetate production is from catalytic, gas-phase partial oxidation of ethylene and acetic acid together.

$$C_2H_4 + CH_3COOH + 1/2O_2 \longrightarrow CH_3COOCH{=}CH_2 + H_2O$$

This oxidation can also be done in the liquid phase. "Syngas", CO and H_2, usually from steam reforming of natural gas, can also be the basis for vinyl chloride. This is a multistep pathway through methanol or methyl acetate made from syngas reacting with syngas.

b) Gross economics. Ethylene and acetic acid oxidation. Basis: 1 kg mol of vinyl acetate.

	Price	Amount	Value
Ethylene	\$0.53/kg	28 kg	\$-14.84
Acetic acid	\$1.00/kg	60 kg	\$-60.00
Oxygen	\$0.04/kg	16 kg	\$- 0.64
Vinyl acetate	\$1.30/kg	86 kg	\$111.80
		Net	\$ 36.32

This result looks attractive, and is the same for both the gas- and liquid phase oxidation. This pathway would be even more favorable if the producer also made the acetic acid by the reaction

$$CH_3OH + CO \rightarrow CH_3COOCH{=}CH_2$$

and made the CO from natural gas or other sources.

Natural gas route to syngas and vinyl acetate. The overall reaction is

$$4CH_4 + 2H_2O \rightarrow CH_3COOCH{=}CH_2 + 7H_2$$

On a basis of 1 kg mol of vinyl acetate, the gross economics is

	Price	Amount	Value
Methane	\$0.13/kg	64 kg	\$ -8.32
H_2O	\$0.008/kg	36 kg	\$ -0.29
Vinyl acetate	\$1.30/kg	86 kg	\$111.80
Hydrogen	\$0.67/kg	14 kg	\$ 9.38
		Net	\$112.57

This pathway looks exceptionally promising. However, because of the numerous steps, and equilibrium and competing reactions it has not proven to be a commercial success as yet.

c) *Ullman's* describes this rather complex process in considerable detail, so the steps of process development will not be shown here. Rather, the reader is referred to *Ullman's*. The oxygen concentration in the ethylene-acetic acid

PROBLEM 4-16 (continued-1)

oxidation reactor feed must be carefully limited to less than 8 volume %. This concentration is only about half of the stoichiometric amount in the equation. This low concentration is required to stay below the lower combustion limit of the gas mixture, primarily for safety reasons. However, it also restricts the oxidation of ethylene. The ethylene conversion is typically on about 10% per pass through the reactor, as excess ethylene is used. As a result of very specific catalysts, the selectivity to of ethylene to vinyl acetate is as high as 94 percent, and 99% based on acetic acid, the more expensive reactant.

Vinyl acetate is a flammable, reactive liquid. It can be stored in steel, aluminum, or stainless steel containers under nitrogen. For shipping the same materials are used, and a stabilizer, such as hydroquinone, is added. The containers must have Symbol F flammable liquid labels.

d) Based on current industrial practice, the ethylene/acetic acid pathway is the most promising when all factors are considered. This is probably due to raw materials costs and, perhaps more importantly, the cost of processing. The combination of processes required for the syngas route is complicated, yields are not so high, and it does not appear to be economically competitive at the present time. The potential of the syngas route is high, as indicated by the preliminary economics, and it is worthy of further attention and development.

e. See problem 4-15 solution for a discussion of using a simulation program to produce a PFD for this problem.

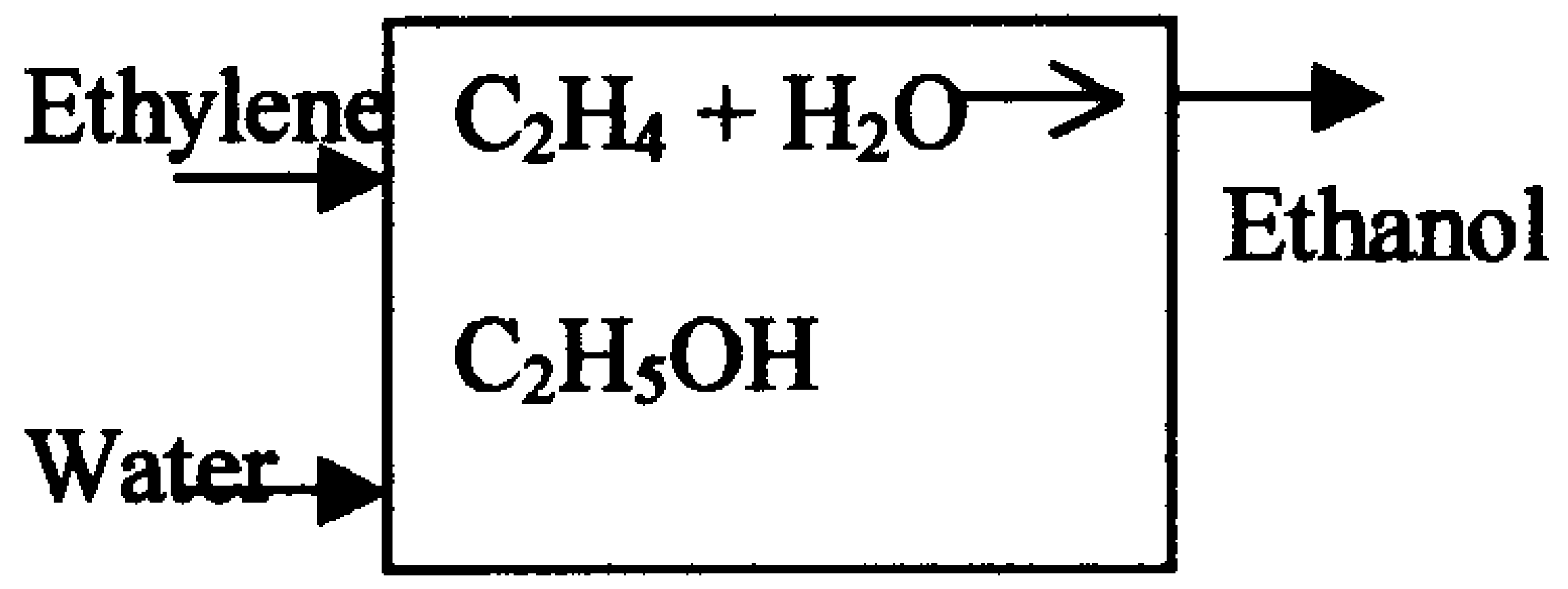
PROBLEM 4-17

Source: *Kirk-Othmer Encyclopedia of Chemical Technology,* 4th ed.

Industrial ethanol can be produced synthetically by reaction of ethylene and water, or by fermentation of a suitable feed. Ethanol and corn prices and data from www.icislor.com and www.ohiocorn.org.

Ethylene pathway. Ethylene and water can be reacted in the liquid or gas phase at elevated temperature and pressure with a catalyst. The gas-phase reaction is used most often.

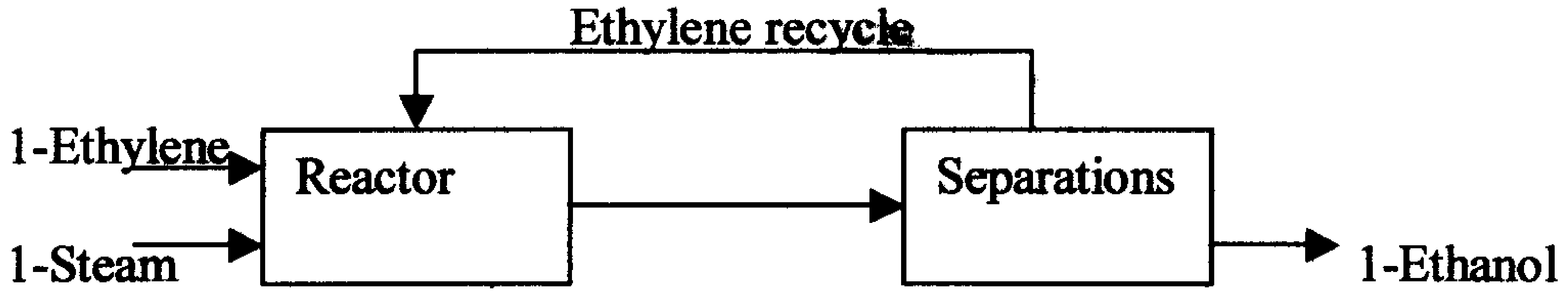
I/O Basis: 1 kg mol ethanol



	Price/ kg	Amount	Cost
C_2H_4	$0.53	28 kg	$-14.84
H_2O	$0.008	18 kg	$ -0.15
C_2H_5OH	$0.84	46 kg	$38.64
		Net	$23.65

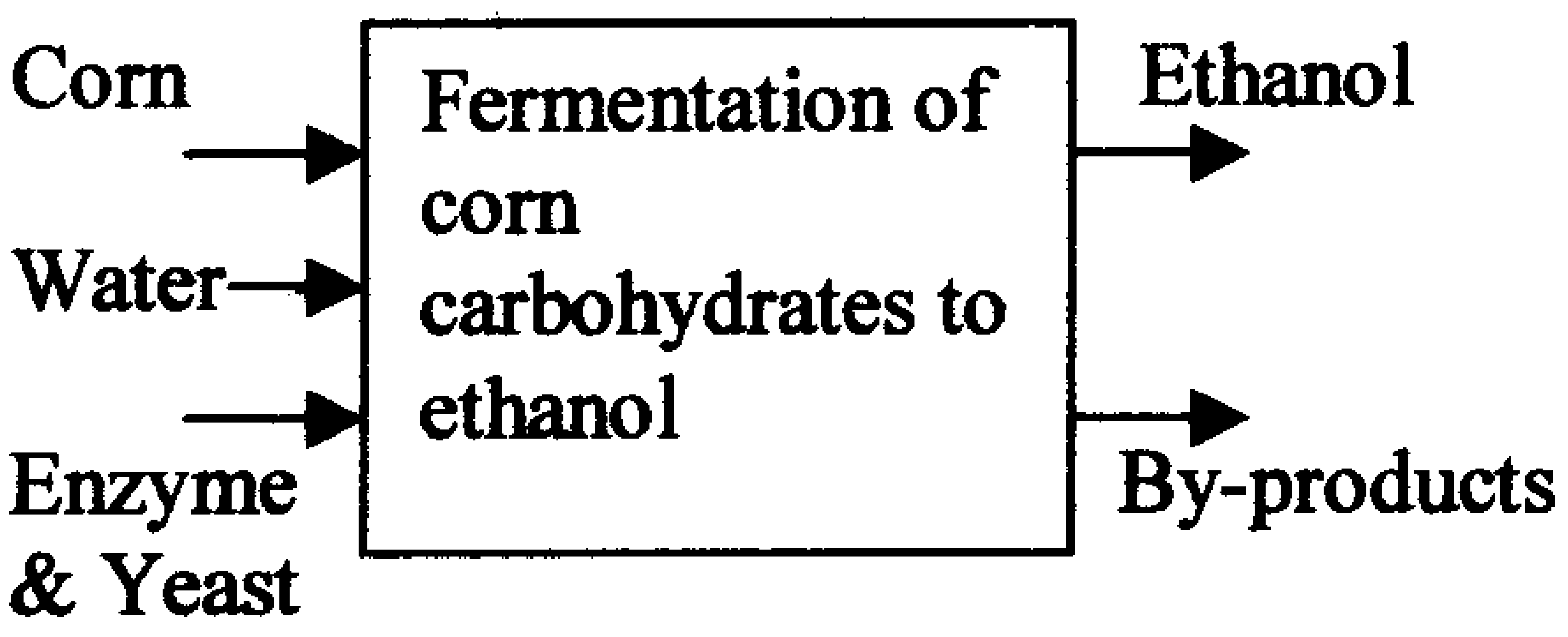
This process looks economically favorable.

Functions



Ethanol from carbohydrate. Ethanol can be produced by fermentation of carbohydrates from a wide variety of sources; fermentation is one of the oldest of human-mediated chemical reactions. In the U.S., corn is the most widely utilized carbohydrate source for this fermentation when industrial/fuel ethanol is to be the product, so this comparison will use corn as the raw material.

I/O Basis: 1 bushel (~56 lb, ~25 kg) corn yields ~2.5 gallons (~7.5 kg of ethanol)



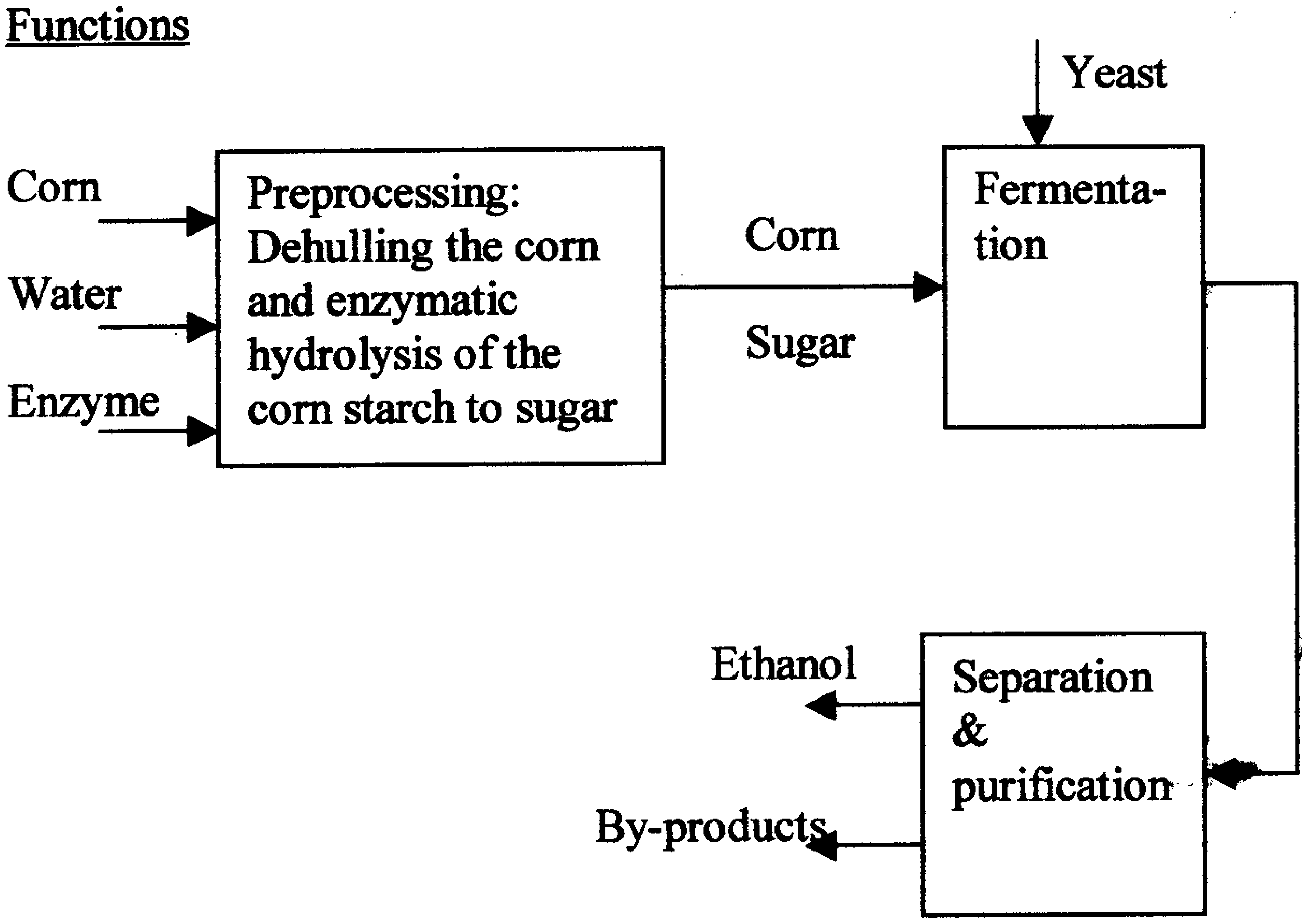
	Price/kg	Amount	Value
Corn	$0.08	25 kg	$-2.00
Ethanol	$0.33	7.5 kg	$ 2.48
		Net	$ 0.48

(~$2.95/kg mol ethanol)

PROBLEM 4-17 (continued-1)

This process looks economically favorable, though not nearly as favorable as the ethylene process. Moreover, the processing is more complex. Several by-products, including animal feed are produced. There is a currently a U.S. government subsidy which helps the operation be profitable. The fermentation step of the process is usually batch, but the rest can be continuous.

Functions

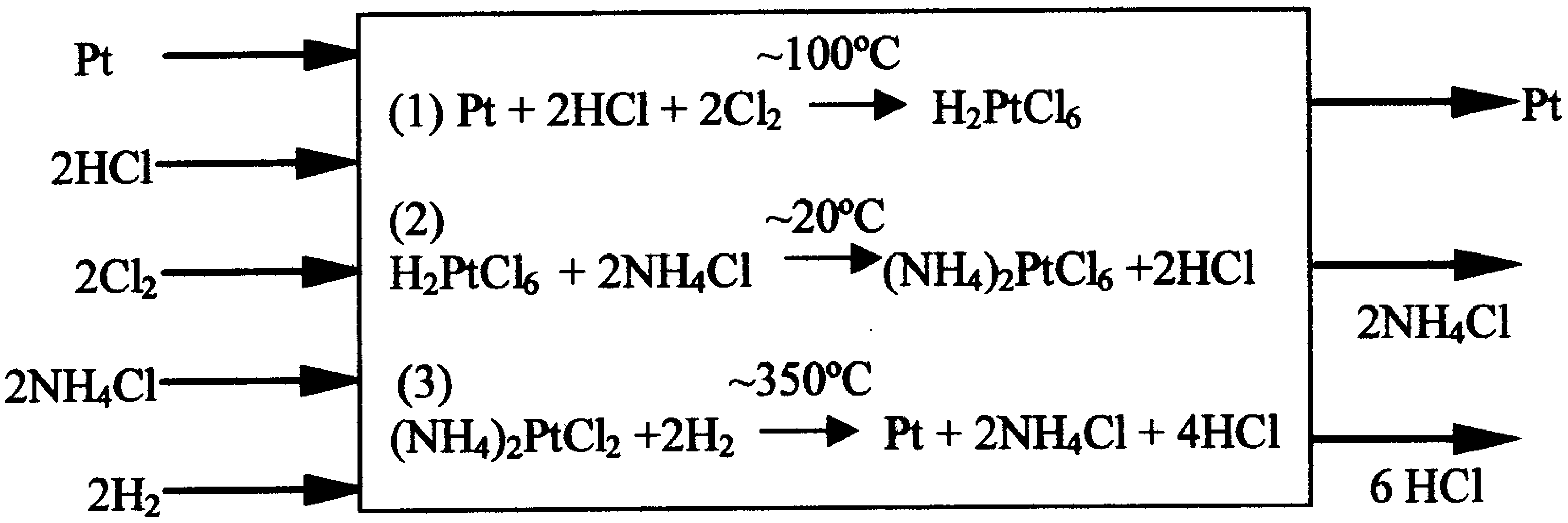


PROBLEM 4-18.

NOTE CORRECTION: There should be 2-g of metal per unit, not 0.5 kg. Sources: *Kirk-Othmer Encyclopedia of Chemical Technology,* 4th ed., and *Ullman's Encyclopedia of Industrial Chemistry.*

Input-output and preliminary economics.

Platinum (and other Pt-group metals, primarily Palladium) can be dissolved from the ceramic, after crushing it, in strong oxidizing acids, such as 6M aqueous hydrochloric acid plus chlorine. The Pt can be precipitated by adding ammonium chloride. The ammonium chloride can be decomposed to Pt powder by heating in the presence of hydrogen.



$$(1)\ Pt + 2HCl + 2Cl_2 \xrightarrow{\sim 100^\circ C} H_2PtCl_6$$

$$(2)\ H_2PtCl_6 + 2NH_4Cl \xrightarrow{\sim 20^\circ C} (NH_4)_2PtCl_6 + 2HCl$$

$$(3)\ (NH_4)_2PtCl_2 + 2H_2 \xrightarrow{\sim 350^\circ C} Pt + 2NH_4Cl + 4HCl$$

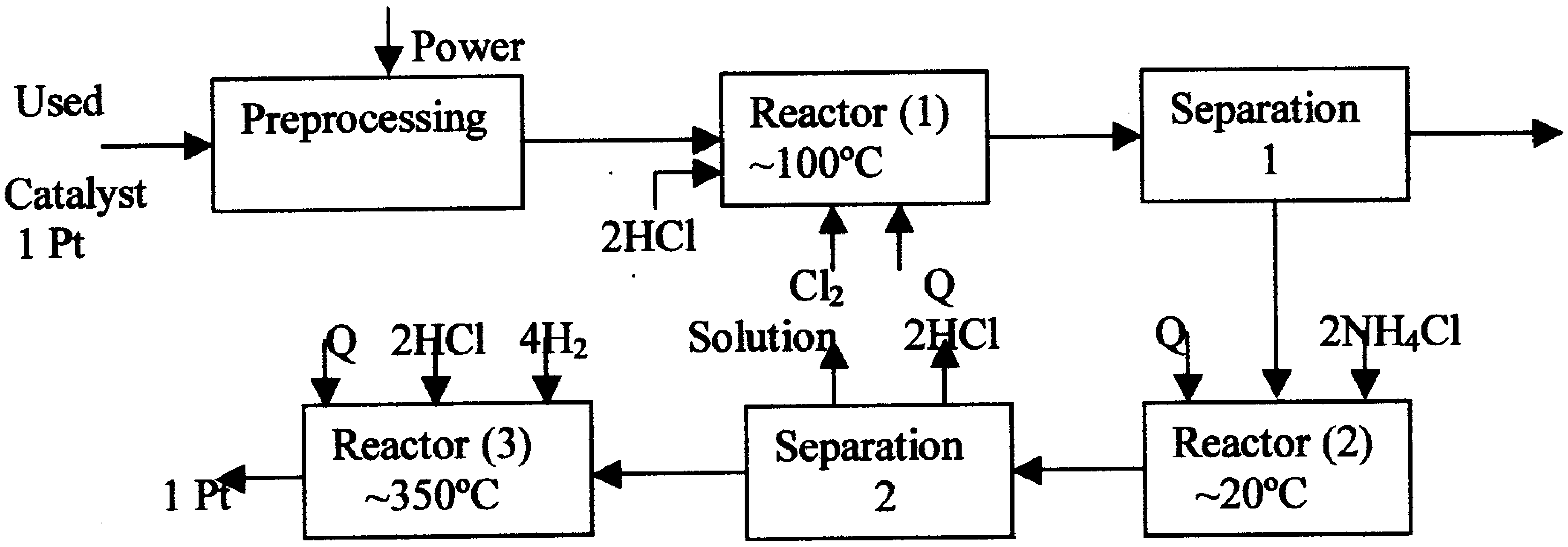
Prices are from *Chemical Market Reporter, Kirk-Othmer, Ullman's,* www.camfordinfo.com, www.chemexpo.com and www.kitco.com, and are time averaged.

Product values: Pt @ \$16/g = 195*16 = \$3,120. NH_4Cl is potentially recoverable, but ignore value. HCl potentially recoverable as HCl and Cl_2, but ignore value

Raw materials' values: Used Pt catalyst on ceramic monolithic is a waste product. However, collecting it for processing may be the most difficult and expensive step of recovery. Nonetheless, it will be considered valueless. For HCl @ \$0.27/kg, value= 0.27*2*36.5 = \$19.7. With Cl_2 @ \$0.21/kg, value = 0.21*2* 71 = \$29.8. Ammonium chloride @ \$0.8/kg has a value of 0.8*2*53.5 = \$85.6. The value of hydrogen @ \$5/kg = 5*2*2 = \$20.0. So the total value of the reactants is \$155.1, almost negligible compared to the potential value of the recovered Pt. However, reprocessing, collection, and other costs have hampered the development of recycling for this material.

PROBLEM 4-18 (continued-1)

Functions diagram. Because of collection difficulties, the capacity at any one site is likely to be small, and because of the difficult and hazardous nature of the processing, it is recommended that the process be batch. A preprocessing step including crushing is needed to prepare steel jacketed catalytic converters for Pt recovery. A reaction (dissolution) step is followed by separation of the crushed ceramic from the solution by filtration. A second reaction step precipitates the Pt as $(NH_4)_2PtCl_6$. The precipitate is filtered from the solution and then heated (calcined) to decompose the solid to Pt metal. The metal is washed. Finishing may be required to obtain powder of the right properties.



PROBLEM 6-1

CORRECTION: The 100 m^2 in the first sentence should be 10 m^2.

Use Eqn. 6-2 with the appropriate exponent.

$$\text{1990 Cost for } 20\,m^2 = \$4{,}200\left(\frac{20}{10}\right)^{0.6} = \underline{\underline{\$6{,}370}}$$

ANS.

$$\text{1990 Cost for } 100\,m^2 = \$6{,}370\left(\frac{40}{20}\right)^{0.6}\left(\frac{100}{40}\right)^{0.81}$$

$$= \$20{,}268.$$

Estimate the 2000 cost using an index from Table 6-2.

Using the Marshall & Swift process-industry installed equipment index,

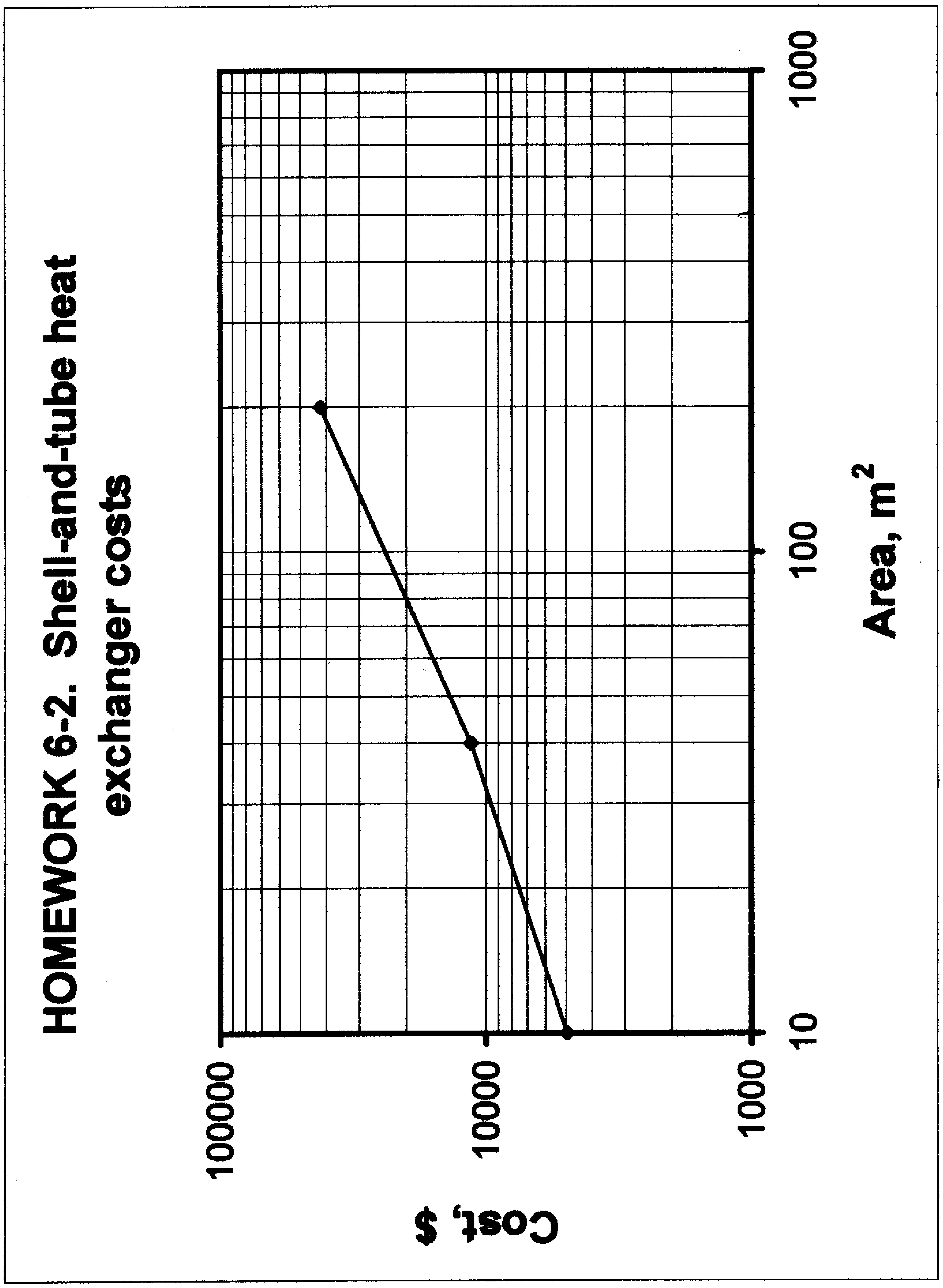
$$\text{2000 cost for } 100\,m^2 = (20{,}268)\left(\frac{1097.7}{929.3}\right) = \underline{\underline{\$23{,}900}}$$

ANS.

For comparison, using the Chemical Engineering Plant Cost index gives

$$\text{2000 cost for } 100\,m^2 = (20{,}268)\left(\frac{394.1}{357.6}\right) = \$22{,}300$$

PROBLEM 6-2



HOMEWORK 6-2. Shell-and-tube heat exchanger costs

PROBLEM 6-3

Use Eq. 6-2 with appropriate exponent. Although it is out of range, in the absence of other information, assume that the 0.93 exponent can be used down to 45,000 kg. Then apply the Marshall & Swift installed equipment cost index from 1990 to 2000 (see Table 6-2 for index values).

$$\text{Installed cost for 300,000 kg tank in 2000} = \$640{,}000\left(\frac{300{,}000}{45{,}000}\right)^{0.93}\left(\frac{1097.7}{929.3}\right)$$

$$= \$\underline{\underline{4.4 \times 10^6}}$$

ANS.

(Note, because of the uncertanties in the method, the value is rounded to 2 digits).

PROBLEM 6-4

Use Eq. 6-2 with appropriate exponent. A value of \$4.4×10⁶ was obtained in problem 6-3, but the unrounded value should be used in the calculations here and the final result rounded.

$$\$4,410276 = 1.1\times10^6\left(\frac{wt.}{135,000}\right)^{0.88}\left(\frac{1097.7}{929.3}\right)$$

$$wt. = (111,030)^{1/0.88} = \underline{\underline{540,000\ kg}}$$

Note that this value lies outside the stated range for the 0.88 exponent. In the absence of other information, assume the extrapolation is reasonable.

PROBLEM 6-5.

Find tank height and surface area.

$\text{VOL.} = 5\,m^3 = \frac{\pi D^2}{4} H \;;\; D = 2\,m,\ H = \text{height}.$

$H = (4)(5)/(2^2\pi) = 1.59\,m$

$\text{Surface Area} = 2\pi D^2/4 + \pi D H$

$= 2\pi(4/4) + 2\pi\,(1.59) = 16.27\,m^2$

The purchased cost of the tank, in 1995 = \$10,900. The installation cost for metal tanks is in the range 30 to 60 %, part of which is for insulation. Select a value, say 40%, without insulation.

1995 installed cost without insulation =

$(1 + 0.4)\$10,900 = \$15,260$

The 1995 insulation cost is

$= \$(40 + 95)/_{m^2}\ (16.27)m^2 = \$2,200.$

Take current = 2000, and use the M&S installed equipment index to update the costs.*

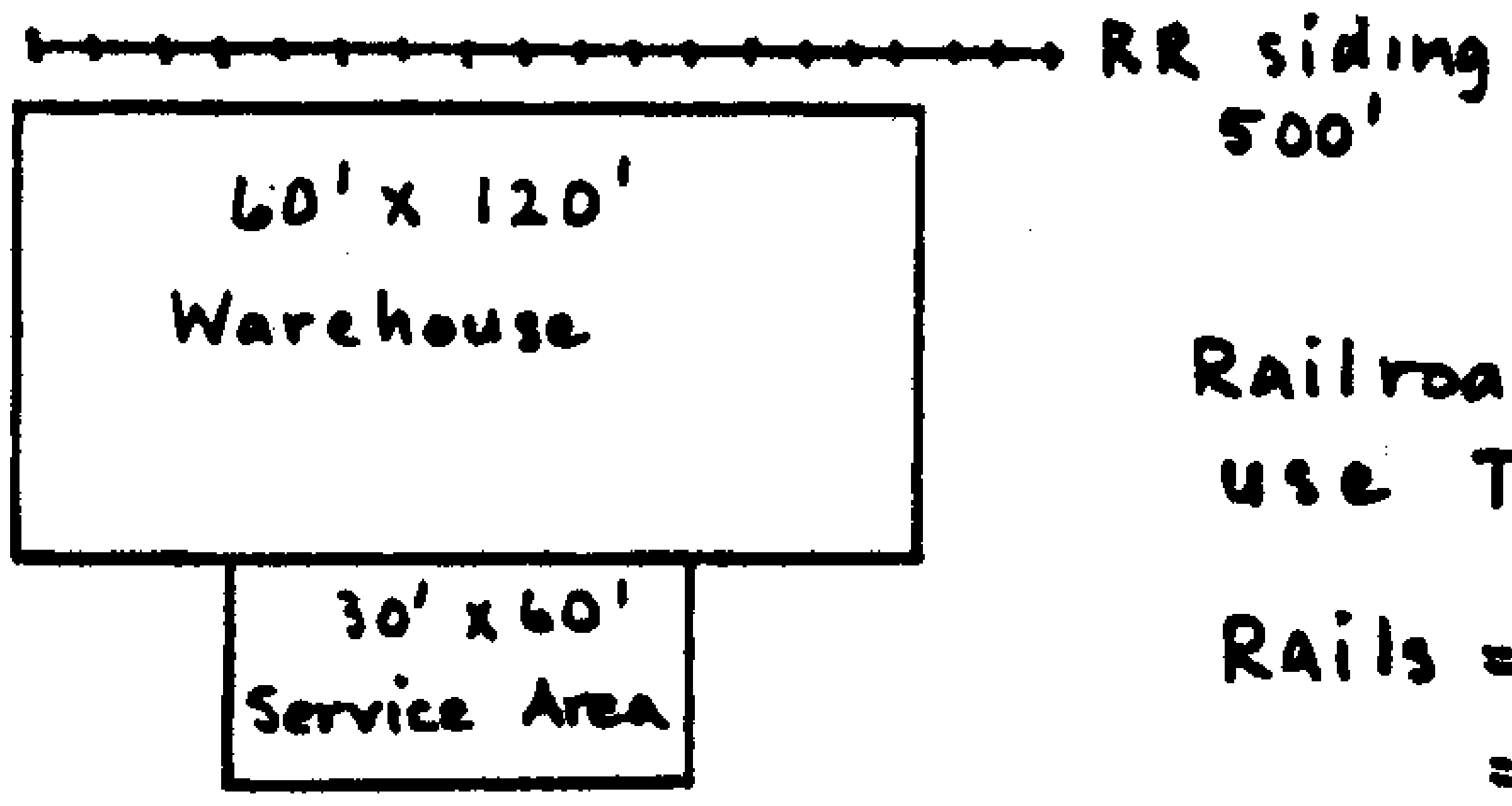
2000 cost of tank installed and insulated $= (15,260 + 2,200)\left(\frac{1097.7}{1029.0}\right)$

$= \underline{\underline{\$\,18,600}}$ ANS.

*It might be preferable to use materials and labor indexes for the insulation if available.

PROBLEM 6-6

The necessary costs can be found in the two tables following this solution (Tables 2 and 3, Appendix B, 4th edition of Peters and Timmerhaus.



Basis: Jan. 1990 cost figures

Railroad siding costs - use Table 3 - Appendix B

Rails = (500 ft)($63/lin.ft) = $31,500 (90 lb Track)

Assume necessity of turnout cost

Turnout cost = $17,030

Total railroad siding cost = $ 48,530

Warehouse costs - use Table 2 - App. B

Steel frame, masonry walls, floor and roof, heating, lighting and plumbing - $28/ft²

Cost = ($28/ft²)(60 ft)(120 ft) = $201,600

(heating was not required, but is included in cost)

Sprinkler costs - use Table 2 - App. B

Dry system, no heat - $2.10/ft²

Cost = ($2.10/ft²)(60ft)(120ft) = $15,120

Asphalt for service area - use Table 3 - App. B, $12.90/yd²

Cost = $12.90/yd²)(60ft/3)(120ft/3) = $10,320

Total cost of proposed addition

48,530 + 201,600 + 15,120 + 10,320 = $ 275,500 <u>Answer</u>

Problem 6-6 (continued)

TABLE 2
Building and construction costs (Jan. 1990)

Item	Unit	Cost, $	Employee—hr to install
Floors			
Asphalt tile	ft^2	1.40	†
Concrete, prestressed, 4-in. thick	ft^2	10.70	†
Steel grating	ft^2	27.81	†
Wood deck, 2-in. thick	ft^2	3.70	†
Foundations, includes excavation, backfill, and forming: flat slab, 1 yd^3 concrete, 5.3 ft^2 forms, 100 lb reinforcing steel‡	yd^3	223	6
Pits and basins: 1 yd^3 concrete, 17.5 ft^2 forms, 115 lb reinforcing steel‡	yd^3	285	8
Walls and piers: 1 yd^3 concrete, 57.0 ft^2 forms, 160 lb reinforcing steel‡	yd^3	593	16
Lumber			
Structural, plain	MBF§	910	30
Structural, creosoted	MBF	1,135	30
Plywood (exterior unsanded):			
$\frac{3}{8}$-in.	ft^2	0.32	1
$\frac{1}{2}$in.	ft^2	0.43	1
$\frac{5}{8}$-in.	ft^2	0.46	1
$\frac{3}{4}$-in.	ft^2	0.54	1
Piling (20–25 ton load, 60 ft long)			
Wood, treated	Ea.	670	†
Wood, untreated	Ea.	570	†
Composite	Ea.	1,180	†
Concrete	Ea.	4,300	†
Test pile	Ea.	8,500	†
Load test	Ea.	10,200	†
Equipment "on and off site"	Per job	14,800	†
Roofs			
Aluminum, corrugated, 0.032-in. thick	ft^2	2.32	†
Built-up, 5-ply	ft^2	0.91	†
Reinforced concrete, 4-in. thick	ft^2	10.70	†
Steel, 26 gauge	ft^2	1.44	†
Transite, $\frac{3}{8}$-in. thick	ft^2	2.47	†
Sprinkler systems, exposed. Add 28% for concealed systems			
Wet system	ft^2	1.65	†
Dry system	ft^2	2.10	†
4-in. alarm valve, wet system	Ea.	1,080	40
4-in. alarm valve, dry system	Ea.	1,650	50
6-in. alarm valve, wet system	Ea.	1,290	40
6-in. alarm valve, dry system	Ea.	1,800	50
Compressor to operate 500 heads (Valves and compressor not included in ft^2 price)	Ea.	980	16
Structural steel			
Grating, $1\frac{1}{4}$-in. standard	ft^2	10.18	0.33

†Labor cost is included in material cost.

‡To adjust: forms, \$3.25/ft^2; reinforcing steel, 44¢/lb.

§MBF refers to 1000 board feet.

(Continued)

Problem 6-6 (Continued)

TABLE 2

Building and construction costs (Jan. 1990) (*Continued*)

Item	Unit	Cost, $	Employee—hr to install
Grating, expanded metal	ft^2	4.80	0.17
Grating, checker plate, $\frac{5}{16}$-in.	ft^2	6.80	0.13
Handrail, standard $1\frac{1}{2}$-in. pipe, 2 rails, welded	Lineal ft	22.60	†
Handrail, bar type	Lineal ft	7.60	0.70
Steel ladder, with safety cage (Add 25% for aluminum)	Lineal ft	76.00	†
Steel ladder, without safety cage (Add 25% for aluminum)	Lineal ft	48.30	†
Operating platforms, including stairs	ft^2	42.40	1.8
Stairway	Vert. ft	91.60	
Stair treads, 12-in. wide, galvanized	Ea.	30.50	1
Building steel, shop fab	lb	0.64	0.01
Platform and support steel, shop flab	lb	0.80	0.02
Toeplate steel, 4 in. × $\frac{1}{4}$-in.	Lineal ft	5.40	†
Walls			
Siding			
Transite	ft^2	2.60	†
Aluminum	ft^2	2.20	†
Steel-coated	ft^2	1.80	†
Brick			
4-in.	ft^2	5.50	†
8-in.	ft^2	9.50	†
10-in.	ft^2	11.60	†
12-in.	ft^2	13.70	†
Concrete block, reinforced			
6-in.	ft^2	5.25	†
8-in.	ft^2	5.50	†
12-in.	ft^2	8.10	†
Windows, industrial			
Steel, fixed	ft^2	16.30	†
Aluminum	ft^2	12.40	†
Wood	ft^2	7.30	†
Total cost of erected buildings (median values)			
Laboratory: steel frame, masonry walls, floor and roof; heating, lighting, and plumbing	ft^2	91	†
Office: steel frame, masonry walls, floor, and roof; heating, lighting, and plumbing	ft^2	62	†
Process building: multilevel, 12-ft clearance, steel platforms, heating, lighting, and plumbing			
Masonry construction	ft^2	41	†
Aluminum on steel	ft^2	43	†
Transite on steel	ft^2	32	†
Open structure: 3-level, steel, with lighting and plumbing	ft^2	29	†
Warehouse: single story, 15-ft clearance. Steel frame, masonry walls, floor, and roof; heating, lighting, and plumbing	ft^2	28	†

†Labor cost is included in material cost.

Problem 6-6 (Continued)

TABLE 2
Building and construction costs (Jan. 1990) (*Continued*)

Item	Unit	Cost, $	Employee—hr to install
Doors			
Metal: Steel frame, 8 × 8 ft, automatic	Ea.	2,085	30
Steel rolling, 12 × 12 ft, manual	Ea.	1,320	24
Swing, with frame, 3 × 7 ft, $1\frac{3}{4}$-in. thick	Ea.	228	†
Wood: Sectional, overhead, 12 × 12 ft	Ea.	1,369	†
Swing exterior, with frame, 3 × 7 ft	Ea.	195	†
Excavation			
Machine	yd^3	4–7	†
Hand	yd^3	23–44	†

†Labor cost is included in material cost.

TABLE 3
Costs for yard improvements (Jan. 1990)

Item	Unit	Cost, $	Employee—hr to install
Docks and wharfs			
All concrete, 100 lineal ft wide	Lineal ft	33,200	†
Timber, with wood deck, 100 lineal ft wide	Lineal ft	10,100	†
Dredging	yd^3	4.20	†
Pipe bridges (includes structural steel and foundations)			
Heavy duty	Lineal ft	65	2.5
Light duty	Lineal ft	41	1.8
Pipe column, steel, extra heavy			
16 ft-4 in. diameter	Ea.	126	1.2
16 ft-6 in. diameter	Ea.	260	1.4
22 ft-6 in. diameter	Ea.	350	1.9
Plant fence			
6 ft chain link (3-strand B.W.)	Lineal ft	7.78	†
3- to 4-ft wide person gate	Ea.	200	†
8-ft wide equipment gate	Ea.	330	†
30-ft wide double gate (manual)	Ea.	664	†
20-ft wide gate (automatic)	Ea.	4,500	†
Relocate plant fence	Lineal ft	5.40	†
Railroads			
Track (90 lb)	Lineal ft	63	†
Track (75 lb)	Lineal ft	48	†
Turnout	Ea.	17,030	†
Ties, creosoted (6 in. × 8 in. × 8 ft)	Ea.	38	†
Grade and ballast	Lineal ft	38.50	†
Cars, ore, 24-in. track			
Capacity, ft^3 Wt, lb			
12 660	Ea.	1,040	†
16 700	Ea.	1,110	†
20 930	Ea.	1,240	†

†Labor cost is included with material cost.

(*Continued*)

Problem 6-6 (continued)

TABLE 3
Cost for yard improvements (Jan. 1990) *(Continued)*

Item	Unit	Cost, $	Employee—hr to install
Locomotives, mine, battery type			
Size, tons Wt, lb			
9 19,000	Ea.	102,000	
12 26,000	Ea.	117,000	
Locomotive, mine, Diesel type			
1½ 3,000 2 cyl	Ea.	31,600	
3 7,000 4 cyl	Ea.	39,400	
Roads and walkways			
Concrete, slab, mesh reinforced			
Thickness, in. Sub-base, in.			
4 6	yd^2	21.50	†
6 6	yd^2	21.80	†
8 6	yd^2	26.50	†
Paving asphalt			
3 12	yd^2	12.90	†
Sewers			
Reinforced concrete, thickness, in. (installed)			
12	Lineal ft	20.90	†
15	Lineal ft	23.30	†
18	Lineal ft	26.10	†
24	Lineal ft	36.00	†
30	Lineal ft	47.40	†
Vitrified clay, thickness, in. (installed)			
4	Lineal ft	14.00	†
6	Lineal ft	15.90	†
8	Lineal ft	19.50	†
12	Lineal ft	30.25	†
15	Lineal ft	38.80	†
18	Lineal ft	48.70	†
24	Lineal ft	71.40	†
Site development			
Clearing and grubbing	yd^2	0.90	†
Grade out	yd^3	2.80	†
Cut, fill, and compact	yd^3	4.50	†
New fill compacted	yd^3	9.00	†
Topsoil	yd^3	14.40	†
Gravel fill	yd^3	15.60	†
Crushed stone (¾ in.)	yd^3	15.60	†
Seeding	yd^2	0.80	†
Sluiceway			
Open, piled, and sheathed	Lineal ft		†
Wells			
200 gpm, 400 ft deep, 15 hp	Ea.	26,000	†
500 gpm, 200 ft deep, 40 hp	Ea.	34,000	†
1200 gpm, 400 ft deep, 75 hp	Ea.	47,000	†

†Labor cost is included with material cost.

PROBLEM 6-7

Use capital investment estimating method C - Percentage of Delivered Equipment Cost, with percentages given for solid processing in Table 6-9.

Purchased equipment cost = \$500,000.

Fixed capital investment = $(3.97)(5\times10^5)$

$= \$\underline{\underline{1.99\times10^6}}$ ANS.

Total capital investment $= (4.67)(5\times10^5)$

$= \$\underline{\underline{2.34\times10^6}}$ ANS.

Engineering and supervision = 33% of purchased equip $= (0.33)(5\times10^5)$

$= \$\underline{\underline{165,000}}$ ANS.

$= \left(\frac{1.65\times10^5}{1.99\times10^6}\right)\times100 = \underline{\underline{8.3\%}}$ of fixed cap. inv.

Contractor's fee = 17% of purchased equip. $= (0.17)(5\times10^5) = \$\underline{\underline{85,000}}$ ANS.

$= \left(\frac{8.5\times10^4}{1.99\times10^6}\right)100 = \underline{\underline{4.3\%}}$ of fixed cap. inv.

PROBLEM 6-8

Table 6-9 shows the cost components as typical percentages of purchased equipment cost, E. However, two of these, buildings and contractor's fee are to change. The problem says that building costs will be high. From Table 6-3 select from the high end of the building range, say 18% of fixed capital investment, FCI.

The total direct plant cost is the sum of the items above it. Thus

$$\text{Direct} = (273/100)E + (18/100)FCI$$

The Contractor's fee is to be 7% of direct costs. The total indirect cost is the sum of the items engineering through contingency. Thus

$$\text{Indirect} = (32+34+4+37)E/100$$
$$+ (0.07)\left[(273/100)E + (18/100)FCI\right]$$

$$FCI = \text{Direct} + \text{Indirect}$$
$$= (2.73)E + (0.18)FCI + 1.07E +$$
$$+ (0.07)((2.73)E + (0.18)FCI)$$

From which

$$FCI - 0.1926\, FCI = 1.197 \times 10^6$$

and

$$FCI = \underline{\$1.483 \times 10^6} \text{ ANS.} \quad (= 4.94E)$$

$$\text{TOTAL DIRECT COST} = \underline{\underline{\$1.086 \times 10^6}} \text{ ANS.} \quad (= 3.62E)$$

$$\text{WORKING CAP.} = (.75)(E) = \$225{,}000$$

$$TCI = FCI + WC = \$\underline{\underline{1.708 \times 10^6}} \quad (= 5.69E)$$

ANS.

PROBLEM 6-9

Using cost and capacity values from Table 6-11 and Eq. 6-11, the turnover ratio, TOR, is

$$TOR = \frac{\text{sales revenue, \$/yr}}{\text{capital investment, \$}} =$$

$$= \frac{(9 \times 10^7 \text{ kg/yr})(86/10^3 \text{ \$/kg})}{\$ 4 \times 10^6} = 1.935 \text{ yr}^{-1}$$

$$\therefore FCI = \frac{(1.3 \times 10^8 \text{ kg/yr})(\$ 86/10^3 \text{ kg})}{1.935 \text{ yr}^{-1}}$$

$$= \underline{\underline{\$ 5.78 \times 10^6}}$$

ANS.

Using the plant capacity ratio Eq. 69, exponent from Table 6-11 = 0.65.

$$FCI = \$ 4 \times 10^6 \left(\frac{1.3 \times 10^8}{9 \times 10^7} \right)^{0.65}$$

$$= \underline{\underline{\$ 5.08 \times 10^6}}$$

ANS.

Chap. 6 - Prob. 10

Use method G in Chapter 6

$$\text{F.C.I.} = \text{total capital investment} - \text{working capital}$$

$$= \$1{,}000{,}000 - 100{,}000 = \$900{,}000$$

$$\text{Turnover ratio} = \frac{\text{gross annual sales}}{\text{fixed capital investment}} = 1.0$$

$$\$/\text{kg selling price} = \frac{\text{fixed capital investment}}{\text{annual production}}$$

$$= \frac{\$900{,}000}{(8000)(365)} = \$0.308 \qquad \underline{\underline{\text{Answer}}}$$

PROBLEM 6-11

Using rate and productivity factors from Table 6-12

$$\text{1990 labor cost in So. Atl.} = \$425{,}000\left(\frac{0.84}{1.06}\right)\left(\frac{0.96}{0.91}\right)$$

$$= \$355{,}300$$

To update this to 1998, use the Engineering News-Record construction index (1967=100 basis)

$$\text{1998 labor cost in So. Atl.} = \$355{,}300\left(\frac{551}{441}\right)$$

$$= \underline{\underline{\$444{,}000}}$$

ANS.

Note, it would be better to use the location and labor specific indexes. However, these are not readily available, so this construction index, of which labor is a major part, is used as a substitute.

PROBLEM 6-12

Basis: 50 kg of 70% soap solution.

Cost delivered to customer, 50kg,

$$= \$20 + (50)(\$1.50/50)$$
$$= \$21.50$$

The 95% solution must result in the same amount of soap, i.e. $(50)(0.7) = 35$ kg, and the total delivered cost.

Amt. of 95% solution $= 35/0.95$
$$= 36.82 \text{ kg}$$

let p = price per 50kg of 95% solution, then $p/50$ = price per kg, and

$$21.5 = 36.82\,(p/50 + 1.5/50)$$

thus

$$p/50 = 21.5/36.82 - 1.5/50$$
$$= 0.554 \ \$/\text{kg}$$

and $p = (50)(0.554) = \underline{\underline{\$27.69/50\text{kg}}}$

ANS.

6.17

PROBLEM 6-13

The main calculations are performed using the spreadsheet for annual total product cost at 100% capacity, **Annual TPC**, from the book's web site.

Because some of the terms in the sum for TPC depend upon TPC, a key to the solution is expressing the total product cost in equation form as:

$$\text{TPC} = \Sigma(\text{all terms not depending on TPC}) + \Sigma(\text{all terms depending on TPC})$$

$$= \Sigma(\text{all terms not depending on TPC}) + \Sigma(\text{Factor}_i)(\text{TPC})$$

which gives,

$$\text{TPC}[1-\Sigma(\text{Factor}_i)] = \Sigma(\text{all terms not depending on TPC})$$

and,

$$\text{TPC} = [\Sigma(\text{all terms not depending on TPC})] / [1-\Sigma(\text{Factor}_i)]$$

The following calculations are made from the spreadsheet results shown on the next page.

a) Manufacturing cost per kg product $= (\$1.244*10^6/\text{y})/(3*10^6\text{kg/y}) = \$0.417/\text{kg}$

b) Total product cost per year $= \$1.427/\text{y}$

c) Profit per kg of product before taxes $=$ gross profit $=$ selling price, \$/kg $-$ TPC/kg

$= \$0.82 - (\$1.427*10^6)/(3*10^6)$

$= \$0.344/\text{kg}$

d) After tax profit per kg product $=$ net profit $=$ (gross profit)$(1-\Phi)$

$= (\$0.344/\text{kg})(1-0.35) = \$0.224/\text{kg}$

6.18

	A	B	C	D	E	F	G	H
1	**Title: Problem 6-13**							
2	**Annual total product cost at 100% capacity**							
3	See Figure 6-7 and 6-8					Date: Nov. 25, 2002		
4	Default, may be changed				Subtotal			
5	Input from problem statement				**RESULT**			
6	Required, may be calculated here, in linked worksheet, or entered manuall							
7	Product:			N/A				
8	Operating time			8000	hours per year			
9	Capacity			3	10^6 kg per year			
10	Fixed Capital Investment, FCI			1.275	million $			
11						User variables		Calculated values, million $
12				Suggested factor		Cost per amount, or factor	Amount, millions	
13	Raw materials					0.09	3	0.270
14	Operating labor					0.08	3	0.240
15	Operating supervision			0.15	of operating labo	0.15	0.240	0.036
16	Utilities					0.05	3	0.150
17	Maintenance and repairs			0.07	of FCI	0.07	1.275	0.089
18	Operating supplies			0.15	of maintenance &	0.15	0.089	0.013
19	Laboratory charges			0.15	of operating labo	0.15	0.240	0.036
20	Royalties (if not on lump-sum basis			0	of TPC	0	1.427	0.000
21	Catalysts and solvents			0		0		0.000
22							**Variable Production Costs =**	**0.835**
23	Taxes (property)			0.02	of FCI	0.02	1.275	0.026
24	Financing (interest)			0	of FCI	0	1.275	0.000
25	Insurance			0.01	of FCI	0.01	1.275	0.013
26	Rent			0	of FCI	0	1.275	0.000
27	Depreciation			0.1	of FCI	0.1	1.275	0.128
28							**Fixed Charges =**	**0.166**
29	Plant overhead, general			0.6	of labor, supervis	0.6	0.365	0.219
30	Plant overhead, packaging					0.008	3	0.024
31							**Plant Overhead =**	**0.243**
32							**Manufacturing cost =**	**1.244**
33	Administration			0.15	of labor, supervis	0.15	0.365	0.055
34	Distribution & selling			0.05	of TPC	0.05	1.427	0.071
35	Research & Development			0.04	of TPC	0.04	1.427	0.057
36							**General Expense =**	**0.183**
37					**TOTAL PRODUCT COST**	**=**	**TPC =**	**1.427**
38								

PROBLEM 6-14

The main calculations are performed using the spreadsheet for annual total product cost at 100% capacity, **Annual TPC**, from the book's web site.

Because some of the terms in the sum for TPC depend upon TPC, a key to the solution is expressing the total product cost in equation form as:

$$\text{TPC} = \Sigma(\text{all terms not depending on TPC}) + \Sigma(\text{all terms depending on TPC})$$

$$= \Sigma(\text{all terms not depending on TPC}) + \Sigma(\text{Factor}_i)(\text{TPC})$$

which gives,

$$\text{TPC}[1-\Sigma(\text{Factor}_i)] = \Sigma(\text{all terms not depending on TPC})$$

and,

$$\text{TPC} = [\Sigma(\text{all terms not depending on TPC})]/[1-\Sigma(\text{Factor}_i)]$$

The following calculations are made from the spreadsheet results shown on the next page.

Manufacturing cost per 100 kg product $= (\$11.122*10^6/\text{y})/(9*10^6\text{kg/y})(100) = \$123.58/100\text{kg}$

6:20

	A	B	C	D	E	F	G	H
1	**Title: Problem 6-14**							
2	**Annual total product cost at 100% capacity**							
3	See Figure 6-7 and 6-8					Date: Nov. 25, 2002		
4	Default, may be changed				Subtotal			
5	Input from problem statement				**RESULT**			
6	Required, may be calculated here, in linked worksheet, or entered manuall							
7	Product:			N/A				
8	Operating time			7200	hours per year			
9	Capacity			9	10^6 kg per year			
10	Fixed Capital Investment, FCI			4	million $			
11						User variables		Calculated values, million $
12				Default factor, user may change		Cost per amount, or factor	Amount, millions	
13	Raw materials					0.25	9	2.250
14	Operating labor					25	0.105	2.628
15	Operating supervision			0.2	of operating labo	0.2	2.628	0.526
16	Utilities							2.765
17	Maintenance and repairs			0.07	of FCI	0.07	4	0.280
18	Operating supplies			0.15	of maintenance &	0.15	0.280	0.042
19	Laboratory charges			0.15	of operating labo	0.15	2.628	0.394
20	Royalties (if not on lump-sum basis			0	of TPC	0	12.788	0.000
21	Catalysts and solvents			0		0		0.000
22					**Variable Production Costs =**			**8.885**
23	Taxes (property)			0.02	of FCI	0.02	4	0.080
24	Financing (interest)			0	of FCI	0	4	0.000
25	Insurance			0.01	of FCI	0.01	4	0.040
26	Rent			0	of FCI	0	4	0.000
27	Depreciation			0.1	of FCI	0.1	4	0.400
28						**Fixed Charges =**		**0.520**
29	Plant overhead, general			0.5	of labor, supervis	0.5	3.434	1.717
30						**Plant Overhead =**		**1.717**
31						**Manufacturing cost =**		**11.122**
32	Administration			0.15	of labor, supervis	0.15	3.434	0.515
33	Distribution & selling			0.05	of TPC	0.05	12.788	0.639
34	Research & Development			0.04	of TPC	0.04	12.788	0.512
35						**General Expense =**		**1.666**
36					**TOTAL PRODUCT COST**	=	**TPC** =	**12.788**
37								

Chap. 6 - Prob. 15

Gross earnings = total sales − direct and indirect production costs
= \$800,000 − (.5)(800,000) − 200,000
= \$200,000

Gross earnings under increased annual sales held constant = \$200,000 = $x - .5x - 1.2\,(200{,}000)$

$x = \$880{,}000$ Answer

If plant operated at full capacity under increased annual sales,

Gross earnings = (1.3)(800,000) − (.5)(1.3)(800,000) − (1.2)(200,000)
= \$280,000

Net profit = (1 − .35)(280,000) = \$182,000 Answer

If total annual sales remained the same as at present,

Gross earnings = 800,000 − 400,000 − (1.2)(200,000)
= 160,000

Net profit = (1 − .35)(160,000) = \$104,000 Answer

If total annual sales decreased to \$700,000

Gross earnings = 700,000 − 350,000 − (1.2)(200,000)
= 110,000

Net profit = (1 − .35)(110,000) = \$71,500 Answer

PROBLEM 6-16

Assume 8000 h/y operating time.

At breakeven, annual values of sales revenue, s, and total product cost, TPC, are equal. Let x = fraction of annual capacity produced at breakeven.

$$s = x\,(5000\ \text{kg/d})(8000/24\ \text{d/y})(1.75\ \$/\text{kg})$$

At breakeven,

$$s = (\$2\times10^6/\text{y})\,x + \$700{,}000/\text{y}$$

setting the two equal and solving gives

$$x = 0.764$$

The production rate is the

$$(0.764)(5000)\left(\frac{8000}{24}\right) = 1.273\times10^6\ \text{kg/y}$$

$$\text{Fixed cost/kg} = \frac{7\times10^5}{1.273\times10^6} = \underline{\underline{\$0.55/\text{kg}}}$$

ANS.

At 100% capacity and \$1.75/kg,

$$\text{Net profit} = (0.65)\left((5000)(8000/24)(1.75) - 2.7\times10^6\right)$$

$$= \$140{,}800/\text{y}$$

If the selling price increases 10%

$$\text{Net profit} = (0.65)\left(1.1(2.917\times10^6) - 2.7\times10^6\right)$$

$$= \$330{,}400/\text{y}$$

And the increase = $\underline{\underline{\$189{,}600/\text{y}}}$ ANS.

PROBLEM 6-17

Fixed capital inv. $= (0.8)(2.5\times10^6)$
$= \$2.0\times10^6$

Annual sales $\simeq (0.5)FCI = \$1.0\times10^6/y$

Annual total product cost
$\simeq \$1.5\times10^6/y$

$\therefore$ There would be an annual loss of $\$0.5\times10^6/y$ ANS.

If a ratio of \$1 of sales per \$1 of fixed capital inv. was achieved, then

Annual sales $\simeq FCI = \$2\times10^6/y$

Gross profit $= 2\times10^6 - 1.5\times10^6$
$= \$0.5\times10^6/y$
$= \dfrac{0.5\times10^6}{2.5\times10^6}\times100 = 20\ \%/y$

and

Net profit $= \dfrac{(0.65)(0.5\times10^6)}{2.5\times10^6}\times100$
$= 13\ \%/y$

Chap. 6 - Prob. 18

Total capital investment = $\$1 \times 10^6$

Total annual sales = $\$1.5 \times 10^6$

Net profits must pay off TCI and R+D in 7 years

Require 12% of sales after taxes = $(.12)(\$1.5 \times 10^6) = \$180{,}000$

34% of R+D is tax free; tax rate is 34% of gross earnings

Basis 1 year

Return after taxes = $180,000

Total R+D cost = R

Net R+D cost = R - am't saved by being tax free
and added to cost of product.

Amount saved by being tax free = R (0.34)

Net R+D cost = R − R(.34) = .66 R

thus

$$180{,}000 = \frac{1.0 \times 10^6 + .66R}{7}$$

$$R = \frac{260{,}000}{.66} = \$394{,}000 \quad \underline{\underline{\text{Answer}}}$$

PROBLEM 6-19

Note that all costs referred to are on a kg basis
Choose as basis, first, 1 kg of product at full capacity

Let
x = total prod. cost/kg of product, full capacity
R = rate of production as kg/yr
T = total prod. cost/kg of product, half capacity

Fixed costs are constant $= (0.35)(10^6)(x)$ in \$/yr
or $= \dfrac{(0.35)(10^6)(x)}{R}$ in \$/kg

Raw materials are constant per kg of product or $= 0.4x$

Other variable costs are directly proportional to $(R)^{1.5}$ or variable costs $= C(R)^{1.5}$ where C is a proportionality constant which may be evaluated when $R = 10^6$ as

$$0.25x = C(10^6)^{1.5} \quad \text{or} \quad C = (0.25)(10^{-9})x$$

thus, the total cost/kg at $R = 0.5 \times 10^6$ kg/yr is

$$T = \frac{(0.35)(10^6)x}{(0.5 \times 10^6)} + 0.4x + (0.25)(10^{-9})(0.5 \times 10^6)^{1.5}x$$

$$= (0.7 + 0.4 + 0.088)x = 1.188x$$

$$\left(\frac{T-x}{x}\right)(100) = \left(\frac{T}{x} - 1\right)100 = (1.188 - 1)(100) = 18.8\%$$

Answer

Note: A common error in this problem, which often occurs, is the assumption that the other variable costs per year are directly proportional to $(R)^{1.5}$.

Chapter 7

7-1. $i = \frac{0.06}{2} = 0.03 \quad N = 10 \cdot 2 = 20 \quad P = \$10{,}000$

$$F = P(1+i)^N$$

$$F = \$10{,}000(1.03)^{20}$$

$$\boxed{F = \$18{,}061}$$

7-2. $F = P(F/P, i, N) = \frac{A\left[(1+i)^N - 1\right]}{i}$

$$\$50{,}000 = \frac{A\left[(1.06)^{10} - 1\right]}{.06}$$

$$\boxed{A = \$3{,}793}$$

$$V = \$50{,}000 - \$5000 = \$45{,}000$$

$$d = \frac{\$45{,}000}{n} = \$4{,}500$$

asset value after 5 years $= \$50{,}000 - 5(\$4{,}500)$

$$= \boxed{\$27{,}500}$$

7-3. (1) <u>End of year</u>

$$F = \frac{A\left[(1+i)^N - 1\right]}{i}$$

$$= \frac{\$10{,}000\left[(1.08)^{10} - 1\right]}{.08}$$

$$= \boxed{\$144{,}866}$$

$$P = F/(1+i)^N$$

$$= \frac{\$144{,}866}{(1.08)^{10}}$$

$$= \boxed{\$67{,}101}$$

(2) Weekly

$A = \$10{,}000/52 = \$192.3077$

$i = .08/52 = 1.538462 \times 10^{-3}$

$N = 520$

$$F = \frac{\$192.3077\left[(1.001538462)^{520} - 1\right]}{1.538462 \times 10^{-3}}$$

$$F = \boxed{\$153{,}022}$$

$$P = \boxed{\$68{,}799}$$

(3) Continuously

$$F = \bar{A}\,(F/\bar{P}, r, N) = \bar{A}\left(\frac{e^{rN} - 1}{r}\right)$$

$$= \$10{,}000\,\frac{e^{(0.08)(10)} - 1}{0.08}$$

$$= \boxed{\$153{,}193}$$

$$P = F/e^{rN}$$

$$= \boxed{\$68{,}834}$$

7-4 $CC = C_V + P$

(1)
$$\left.\begin{aligned} F &= P(1+i)^N \\ F - P &= C_R \end{aligned}\right\} \text{Combine}$$

$$P\left[(1+i)^N - 1\right] = C_R$$

$$\boxed{CC = C_V + \frac{C_R}{(1+i)^N - 1}}$$

(2)
$$\left.\begin{aligned} F &= Pe^{rN} \\ F - P &= C_R \end{aligned}\right\} \text{Combine}$$

$$P\left[e^{rN} - 1\right] = C_R$$

$$\boxed{CC = C_V + \frac{C_R}{e^{rN} - 1}}$$

7-5

(1) $C_V = \$20{,}000$ 6-year life $i = 0.06$

(2) $C_V = \$34{,}000$ $C_S = \$4000$ 10-year life

$$CC_1 = \$20{,}000 + \frac{\$20{,}000}{(1.06)^6 - 1}$$

$$= \underline{\$67{,}788}$$

$$CC_2 = \$34{,}000 + \frac{(\$34{,}000 - \$4{,}000)}{(1.06)^{10} - 1}$$

$$= \underline{\$71{,}934}$$

$$\boxed{\text{Heat exchanger (1) is cheaper}}$$

7-6. (1) $C_v = \$10{,}000$ 10-year life $i = 0.06$

(2) $C_v = ?$ (repair) 3-year life

$$CC_1 = \$10{,}000 + \frac{\$10{,}000}{(1.06)^{10}-1}$$

$$CC_2 = C_{repair} + \frac{C_{repair}}{(1.06)^3-1}$$

Set $CC_1 = CC_2$ & solve for C_{repair}

$$\boxed{C_{repair} = \$3{,}632}$$

7-7. Stock dividends are not deductible

\$5,000,000 @ 8% → \$400,000 / yr in dividends

Bond interest payments are deductible

\$5,000,000 @ 6% → \$300,000/ yr in interest

@ 35% → \$105,000 is deducted

actual cost \$195,000 / yr

$$\boxed{\text{Difference } \$205{,}000/\text{yr}}$$

7-8 (1) \$1,000,000 investment with annual cash flow of \$200,000

(2) \$600,000 investment with annual cash flow of \$220,000

+ \$400,000 @ 6% = \$24,000 in interest.

Option (2) is better just from the cash flow difference. The added interest income makes it even better.

7-9 $FCI = \$20 \times 10^6$

$Tax_{prop} = 1\%$ of FCI

$Tax_{state} = 5\%$ of Gross Earnings (GE)

$Tax_{Fed} = 35\%$ "

$$GE - 0.05GE - 0.35GE - 0.01 \cdot FCI = \$2{,}000{,}000$$

$$GE = \$3{,}000{,}000$$

2nd plant

$Tax_{prop} = 0.04 \cdot FCI$

$Tax_{state} = 0.02 \cdot GE$

$$\text{Net Inc.} = (\$3{,}000{,}000)(1-.35-.02) - 0.04(20 \times 10^6)$$

$$\boxed{= \$1{,}090{,}000}$$

7-10. $F = \$50{,}000$ $i = 0.06$

$N = 10$

$A = \$400/yr$

$$F = \frac{A[(1+i)^N - 1]}{i} + P(1+i)^N \quad \text{solve for } P$$

$$\$50{,}000 = \frac{\$400[1.06^{10}-1]}{.06} + P(1.06)^{10}$$

$$\boxed{P = \$24{,}976}$$

7-11 MACRS - Cap. Gain $= \$7000 - \$1728 = \$5272$

$\text{Tax}_{CG} = .20(\$5272) = \$1054.40$

SL - Cap Gain $= \$7000 - \$2000 = \$5{,}000$

$\text{Tax}_{CG} = \$1000$

Income Tax : $0.35(\$8272 - 8000) = \$95.20$

Tax paid w/ SL over that paid w/ MACRS

DIFFERENCE: $\$95.20 - 54.40 = \boxed{\$40.80}$

7-12 (1) Depreciation = \$200,000/yr

Taxable income = \$200,000 − \$200,000 = 0

Cash flow = \$200,000/yr

(2) Depreciation = \$120,000/yr

Taxable income = \$220,000 − \$120,000 = \$100,000

Taxes = \$35,000

Cash Flow = \$220,000 − \$35,000 = \$185,000/yr

Interest income is \$24,000/yr (see 7-8)

Total cash flow = \$209,000/yr

Option (2) is better

7-13 SL: \$10,000/yr

\$20,000 paid in 2 yrs (40%)

MACRS: 5th year 0.1152(\$50,000) = \$5760

20% + 32% in 2yrs = 52%

7-14. 1st year of MACRS - 20%

20% of $850,000 = $170,000

SL: $d = \frac{\$850{,}000}{9.5} = \$89{,}474$ /yr

Difference = $80,526 less deducted in SL

Taxes = .35($80,526)

= $\boxed{\$28{,}184}$ less paid w/ MACRS

7-15

$6,000,000 5yr (20%/yr) = $1,200,000/yr

$4,000,000 10yr (10%/yr) = $400,000/yr

Total = $1,600,000
1st year

MACRS (15yrs) → 5% in 1st year = $500,000

7-16 unit of production method $10/unit

year	units	deprec.	cum
1	200	$2000	$2000
2	400	$4000	$6000
3	800	$8000	$14,000
4	1600	$16,000	$30,000
5	1600	$16,000	$46,000
6	.	.	.
7	:	:	:
8	.	.	.
9	1600	$16,000	$110,000

SL w/ 9 yrs $d = \frac{\$110{,}000}{9} = \boxed{\$12{,}222/\text{yr}}$

7-17

SL: $d = \dfrac{\$35{,}000 - \$5{,}000}{5} = \$6000$

MACRS: percentages from Table 7-9

sum-of-the-digits: $d = \dfrac{2(n-a+1)}{n(n+1)}(V - V_s)$

method/Depr.

Year	SL	MACRS	SOTD
1	\$6,000	\$6,000	\$10,000
2	"	\$9,600	\$8,000
3	"	\$5,760	\$6,000
4	"	\$3,456	\$4,000
5	"	\$3,456	\$2,000
6	0	\$1,728	0

7-18

$$d = \frac{2(n-a+1)}{n(n+1)}(V - V_s)$$

where n is the recovery period

a is the year for which the depreciation allowance, d, is being determined

V is the original equipment cost

V_s is the salvage value

$d = \$2381$

$a = 2$

$V = \$10{,}000$

$V_s = 0$

solve for n

$\boxed{n = 6 \text{ years}}$

CHAPTER 8

Chap. 8 - Prob 1:

$$F = P\left(1+\frac{i}{m}\right)^{Nm} \qquad (\text{eq. } 7\text{-}6)$$

(a) From the problem statement

$P = \$10,000$

$i = 0.06$

$N = 10$

$m = 12$

$$\therefore \quad F = 10,000\left(1+\frac{0.06}{12}\right)^{120}$$

$$= 18,193.97$$

Total amount of funds = \$18,193.97 Answer

(b) $P = \$10,000$

$F = \$20,000$

$i = 0.06$

$m = 2$

To calculate N

$$20,000 = 10,000\left(1+\frac{0.06}{2}\right)^{2N}$$

$\Rightarrow$ $N = 11.7$ years Answer

(c) $$F = Pe^{rN} \qquad (\text{eq. } 7\text{-}18)$$

$F = 2P$ and $r = 0.06$

Substituting

$$2P = Pe^{0.06N}$$

$$\Rightarrow \quad N = \frac{\ln 2}{0.06}$$

$\Rightarrow$ $N = 11.55$ years Answer

Chap. 8 – Prob 2:

Total investment = T.I. = Working capital + fixed capital

$$T.I. = 0.25\,(T.I.) + 10\times10^6$$

$$\Rightarrow \quad T.I. = \$\,13.33\times10^6$$

$$\text{Percent return on investment} = \frac{\text{Annual profit}}{\text{Total investment}}\times 100$$

$$= \frac{3\times10^6}{13.33\times10^6}\times 100$$

$$= 22.5$$

Answer

$$\text{Payout period} = \frac{\text{Fixed capital investment}}{\text{Annual profit} + \text{Annual depreciation}}$$

$$= \frac{10\times10^6}{3\times10^6 + 1\times10^6}$$

$$= 2.5 \text{ years}$$

Answer

Chap. 8 – Problem 3

Let the total capital investment be T

Then, fixed capital investment = 0.85 T

$$PBP = \frac{\text{Fixed capital investment}}{\text{Annual cash flow}}$$

$$\Rightarrow \text{Annual cash flow} = A_j = \frac{0.85T}{PBP}$$

$$= \frac{0.85T}{5}$$

$$= 0.17T$$

Net profit = Cash flow + depreciation

Depreciation = 10% of fixed capital investment

$$= (0.1)(0.85T)$$

$$= 0.085T$$

$$\therefore \text{Net Profit} = A_j + 0.085T$$

$$= (0.17 + 0.085)T$$

$$= 0.255T$$

$$ROI = \left(\frac{\text{Net profit}}{\text{Total Capital Investment}}\right) \times 100$$

$$= \frac{0.255T}{T} \times 100$$

$$= 25.5\%$$

Answer

Chap. 8 – Prob. 4

Cost of replacing Pump A after 4 years = C_R = \$20,000 – \$2,000

= \$18,000

Installed cost of Pump A = C_V = \$ 20,000

Annual effective interest rate = r = 0.15

Number of years of useful life = n = 4 years

$$\text{Capitalized cost} = K = C_V + \frac{C_R}{(1+i)^n - 1}$$

$$= 20{,}000 + \frac{18{,}000}{(1+0.15)^4 - 1}$$

$$= \$44{,}031.84$$

To be competitive, the capitalized cost of Pump B should be the same as Pump A, i.e., \$44,031.84

Capitalized cost of Pump B = K = \$44,031.84

Cost of replacing Pump B after n years = C_R = 25,000 – 4000

= \$ 21,000

Installed cost of Pump B = C_V = 25,000

Annual effective interest rate = r = 0.15

Using $K = C_V + \dfrac{C_R}{(1+i)^n - 1}$

we have

$$44{,}031.84 = 25{,}000 + \frac{21{,}000}{(1+0.15)^n - 1}$$

$\Rightarrow$ n = 5.3 years

Service life of Pump B = 5.3 years **Answer**

Chap. 8 — Prob. 5

Basis : 1 year

For 0.025 m insulation

$$\text{Money saved on heat} = \frac{(88)(3600)(24)(300)(1.50)}{1,000,000} = \$3,421.44$$

$$\text{Fixed charges} = (0.1)(8,000) = \$800$$

$$\text{Total savings} = 3421.44 - 800 = \$2621.44$$

For 0.051 m insulation

$$\text{Money saved on heat} = \frac{(102)(3600)(24)(300)(1.50)}{1,000,000} = \$3965.76$$

$$\text{Fixed charges} = (0.1)(10,100) = \$1,010$$

$$\text{Total savings} = 3,965.76 - 1,010 = \$2955.76$$

For 0.076 m insulation

$$\text{Money saved on heat} = \frac{(108)(3600)(24)(300)(1.50)}{1,000,000} = \$4199.04$$

$$\text{Fixed charges} = (0.1)(11,100) = \$1110$$

$$\text{Total savings} = 4199.04 - 1110 = \$3,089.04$$

For 0.102 m insulation

$$\text{Money saved on heat} = \frac{(111)(3600)(24)(300)(1.50)}{1,000,000} = \$4315.68$$

$$\text{Fixed charges} = (0.1)(11,500) = \$1150$$

$$\text{Total savings} = 4315.68 - 1150 = \$3165.68$$

Comparing 0.025 m insulation to 0.051 m insulation

$$\text{Return on investment} = \left(\frac{2955.76 - 2621.44}{10,100 - 8000}\right)(100) = 15.92\% \Rightarrow \underline{\text{Prefer } 0.051}$$

Comparing 0.051 m insulation with 0.076 m insulation

$$\text{Return on investment} = \frac{(3089.04 - 2955.76)}{(11,100 - 10,100)}(100) = 13.33\% \Rightarrow \underline{\text{Prefer } 0.051}$$

Comparing 0.051 m insulation with 0.102 m insulation

$$\text{Return on investment} = \frac{(3165.68 - 2955.76)}{(11500 - 10100)}(100) = 14.99\% \Rightarrow \underline{\text{Prefer } 0.051}$$

Answer

Chap. 8 – Prob. 6

Cost of replacing Pump A after 2 years $= C_{RA} = \$15{,}000$
Installed cost of Pump A $= C_{VA} = \$15{,}000$
Service life of Pump A $= n_A = 2$ years

Capitalized cost of Pump A $= K_A = C_{VA} + \dfrac{C_{RA}}{(1+i)^{n_A}-1}$

$$= 15{,}000 + \frac{15{,}000}{(1+i)^2-1}$$

Cost of replacing Pump B after 5 years $= C_{R_B} = \$22{,}000$
Installed cost of Pump B $= C_{VB} = \$22{,}000$
Service life of Pump B $= n_B = 5$ years

Capitalized cost of Pump B $= K_B = 22{,}000 + \dfrac{22{,}000}{(1+i)^5-1}$

To be competitive, $K_A = K_B$

$$\Rightarrow \quad 15{,}000 + \frac{15{,}000}{(1+i)^2-1} = 22{,}000 + \frac{22{,}000}{(1+i)^5-1}$$

Solving by trial and error, $i = 0.63$

Annual interest rate for the two pumps to be competitive is 63%. **Answer**

The pumps are competitive at a very high interest rate. Suppose we assume a more reasonable rate of 10%.

Then, the capitalized cost of Pump A $= 15{,}000 + \dfrac{15{,}000}{(1+0.1)^2-1}$

$$= \$86{,}428.57$$

The capitalized cost of Pump B $= 22{,}000 + \dfrac{22{,}000}{(1+0.1)^2-1}$

$$= \$58{,}035.44$$

I would recommend Pump B as significantly less money needs to be invested to replace it. **Answer**

Chap 8 – Prob. 7

15% return is required for the design to be acceptable.

Comparing No. 2 to No. 1, incremental return is:

$$\left(\frac{3{,}000 - 2{,}800}{12{,}000 - 10{,}000}\right) \times (100) = 10\%. \qquad \text{Reject design No. 2}$$

Comparing No. 3 to No. 1, incremental return is:

$$\left(\frac{3{,}000 - 2{,}350}{14{,}000 - 10{,}000}\right) \times (100) = 16.25\%. \qquad \text{Reject design No. 1}$$

Comparing No. 3 to No. 4, incremental return is:

$$\left(\frac{2{,}350 - 2{,}100}{16{,}000 - 14{,}000}\right) \times (100) = 12.5\%. \qquad \text{Reject design No. 4}$$

Based on the above analysis, Design No. 3 is recommended

Answer

Chap. 8 – Prob. 8

(a) From Table 8-3, we have:

All expenses = cash expenses + depreciation

Cash expenses = \$2,000,000

Depreciation = \$1,000,000 (assuming straight line depreciation of total capital over 10 years)

All expenses = 2,000,000 + 1,000,000
= \$3,000,000

Income tax = (revenue – all expenses) (tax rate)
= (\$8,000,000 – \$3000000) (0.35) (Assume a tax rate of 35%)

= \$1,750,000

Net profits = revenue – all expenses – income tax
= \$8,000,000 – \$3,000,000 — \$1,750,000
= \$3,250,000

Total capital investment = \$10,000,000

$$ROI = \frac{\text{Net profits}}{\text{Total capital investment}}$$

$$= \frac{3{,}250{,}000}{10{,}000{,}000}$$

= 0.325 or 32.5% <u>Answer</u>

b) Cash flow A_j = net profits + depreciation (Table 8-3)
= \$3,250,000 + \$1,000,000
= \$4,250,000

Fixed capital investment = Total capital investment – working capital
= \$10,000,000 — \$1,000,000
= \$9,000,000

$$PBP = \frac{\$9{,}000{,}000}{4{,}250{,}000} = 2.11 \text{ years}$$ <u>Answer</u>

Chap. 8 – Prob. 9

Insurance cost without sprinklers:

$(0.5 \times 10^6)(0.011)(20) = 0.11 \times 10^6$

$(0.4 \times 10^6)(0.0095)(20) = \underline{0.076 \times 10^6}$

Total = \$186,000

Insurance cost with sprinklers:

$(0.52 \times 10^6)(0.011)(20)(3/4) = 0.086 \times 10^6$

$(0.4 \times 10^6)(0.0095)(20)(3/4) = \underline{0.057 \times 10^6}$

Total = \$143,000

Installation + maintenance costs = 20,000 + (300)(20)
= \$26,000

Total cost with sprinklers = 143,000 + 26,000
= \$169,000

Savings in 20 years with sprinkler system = \$17,000

Annual % return = $\frac{17,000}{(20)(20,000)} \times 100 = 4.25\%$

This is less than the present 8% return; therefore, from the viewpoint of the stockholders, the sprinkler system <u>should not be purchased.</u> From the viewpoint of safety, the 4.25% return on investment might be considered acceptable.

Chap 8 - Prob 10

(a) Annual cash flow = Annual revenue − annual operating expense + annual depreciation

We first compute the amount depreciated using \$50 million as the basis and an MACRS depreciation schedule with a class life of 5 years (Table 7-9)

Year	Depreciation rate, %	Amount depreciated, in millions of \$
1	20.00	10.00
2	32.00	16.00
3	19.20	9.60
4	11.52	5.76
5	11.52	5.76
6	5.76	2.88
7	0.00	0.00

Based on the amount depreciated, the annual cash flow can be computed, as shown below (all figures, except year, in millions of dollars)

Year	Annual revenue	Annual operating exp.	Amount dep.	Cash flow
1	7.0	4.0	10.00	13.00
2	10.0	5.6	16.00	20.40
3	15.0	6.8	9.60	17.80
4	20.0	7.8	5.76	17.96
5	22.5	8.8	5.76	19.46
6	24.0	9.6	2.88	17.28
7	25.0	10.0	0.00	15.00

Answer

(b) Net present worth = $\sum_{j=1}^{N} PWF_{cf,j} [s_j - c_j - d_j](1-\phi) + rec_j + d_j] - \sum_{j=-b}^{N} PWF_{v,j} F_j$

Assume that the tax rate = $\phi = 0.35$

Assume that the recovery of working capital = $rec_j = 0$

$PWF_{cf,j} = (1 + 0.15)^{-j}$

s_j = annual revenue from the above table

c_j = annual operating expense from the above table

d_j = annual depreciation from the above table

Chap 8 - Prob 10 (Cont'd)

Substituting, we get:

$$\begin{aligned}\text{Net present worth} = {} & (1.15)^{-1}\{(7\times10^6 - 4\times10^6 - 10\times10^6)(1-0.35) + 10\times10^6\} \\ & + (1.15)^{-2}\{(10\times10^6 - 5.6\times10^6 - 16\times10^6)(1-0.35) + 16\times10^6\} \\ & + (1.15)^{-3}\{(15\times10^6 - 6.8\times10^6 - 9.6\times10^6)(1-0.35) + 9.6\times10^6\} \\ & + (1.15)^{-4}\{(20\times10^6 - 7.8\times10^6 - 5.76\times10^6)(1-0.35) + 5.76\times10^6\} \\ & + (1.15)^{-5}\{(22.5\times10^6 - 8.8\times10^6 - 5.76\times10^6)(1-0.35) + 5.76\times10^6\} \\ & + (1.15)^{-6}\{24\times10^6 - 9.6\times10^6 - 2.88\times10^6)(1-0.35) + 2.88\times10^6\} \\ & + (1.15)^{-7}\{25\times10^6 - 10\times10^6 - 0)(1-0.35) + 0\} \\ & - 50\times10^6 \\ = {} & \$ -13{,}885{,}994.52\end{aligned}$$

Answer

(c) DCFR

$$\begin{aligned}0 = {} & (1+r)^{-1}\{5{,}450{,}000\} + (1+r)^{-2}\{8{,}460{,}000\} + (1+r)^{-3}\{8{,}690{,}000\} \\ & + (1+r)^{-4}\{9{,}946{,}000\} + (1+r)^{-5}\{10{,}921{,}000\} + (1+r)^{-6}\{10{,}368{,}000\} \\ & + (1+r)^{-7}\{9{,}750{,}000\} - 50{,}000{,}000\end{aligned}$$

Calculating by trial and error

$r \approx 0.035$ or 3.5%

Answer

Chap 8 - Prob. 11

Annual savings by choice of steam unit

Fuel	\$70,000
Maintenance	\$ 3,000
Insurance and taxes	− \$ 6,000
Depreciation : $\frac{400,000}{10} - \frac{600,000}{20}$	= \$10,000
Total savings	= \$ 77,000

% incremental return on investment if steam unit chosen :

$$\left(\frac{77,000}{600,000 - 400,000}\right)(100) = 38.5\%$$

12% return on investment is required ; thus the steam turbine unit is an acceptable alternative.

Answer

Chap 8 – Prob 12

Let x be the initial manufacturing cost at both locations

For old plant:

Capital investment = \$130,000

Manufacturing cost = \$150,000 + 60,000 + x

= \$ 210,000 + x

For new plant:

Capital investment = \$200,000

Manufacturing cost = \$120,000 + 70,000 + (0.02)(200,000) + x + depreciation

= \$ 194,000 + x + depreciation

A return of 9% after taxes of 35% are required

This is equivalent to $\frac{9}{1-0.35}$ = 13.8% return before taxes

$$\text{Rate of return} = 0.138 = \frac{-\Delta \text{ manufacturing costs}}{\Delta \text{ investment}}$$

$$= \frac{\text{New manufacturing cost} - \text{old manufacturing cost}}{\text{Old investment} - \text{new investment}}$$

$$= \frac{(194{,}000 + x + \text{depreciation}) - (210{,}000 + x)}{130{,}000 - 200{,}000}$$

$$= \frac{\text{depreciation} - 16000}{-70{,}000}$$

$\Rightarrow$ depreciation = (0.138)(−70,000) + 16,000

= \$ 6340 per year

$$\text{Recovery period for new investment} = \frac{\text{Capital investment}}{\text{annual depreciation}}$$

$$= \frac{200{,}000}{6340}$$

= 31.5 years **Answer**

Chap 8 - Prob 13

Definitions from Table 8-3 are used in this solution

Old unit:

Let the revenue be = R

Let the cash expense be = E

Annual depreciation allowed = $\frac{40,000}{10} = \$4,000$ (assuming straight line method)
(Note that in this method no salvage value may be taken as stated on page 311)

$\therefore$ All expenses $= E + 4000$

Tax $= T =$ (revenue − all expenses) (tax rate)
$= (R - E - 4000)(0.35)$

Net Profit $= P_{old} =$ Revenue − All expenses − Tax
$= R - (E + 4000) - (R - E - 4000)(0.35)$
$= 0.65(R - E) - 2600$

New Unit:

Revenue $= R$ (same as old unit)

Cash expenses $= E - 15,000$

Annual depreciation allowed $= \frac{70,000}{10} = \$7,000$

All expenses $= E - 15,000 + 7000 = E - 8,000$

Tax = (revenue − all expenses) (tax rate)
$= (R - E + 8000)(0.35)$

Net profit $= P_{new} =$ Revenue − All expenses − Tax
$= R - (E - 8000) - (R - E + 8000)(0.35)$
$= 0.65(R - E) + 5200$

Increase in profit if one switches to new unit $= P_{new} - P_{old}$
$= \$7,800$

New investment $= 70,000 - 5000 = \$65,000$

% Return $= \frac{7800}{65000} \times 100 = 12\%$. Since this is less than 15% do NOT make change.

Chap 8 - Prob 14

Fixed capital investment = \$1,000,000

Depreciation = $\frac{1{,}000{,}000}{5}$ = \$ 200,000

Cash flow = $(s_j - c_j - d_j)(1-\phi) + rec_j + d_j$ where $\phi = 0.35$

Present worth = $PWF_{c,f,j} \{ \text{cash flow} \}$

where $PWF_{c,f,j} = \left(\frac{e^r - 1}{r}\right) e^{-rj}$ where $r = 0.10$

Using the above formulae, we can generate the following spreadsheet

Year	Revenue	Product Cost	Depreciation	Cash flow	Present Worth
1	500,000	100,000	200,000	330,000	314,036.52
2	500,000	100,000	200,000	330,000	284,151.99
3	500,000	100,000	200,000	330,000	257,111.36
4	500,000	100,000	200,000	330,000	232,643.98
5	500,000	100,000	200,000	330,000	210,504.97

If we assume an inflation rate of 5% for both revenues and expenses, we get the following:

Year	Revenue	Product Cost	Depreciation	Cash flow	Present Worth
1	500,000	100,000	200,000	330,000	314,036.52
2	525,000	105,000	200,000	343,000	295,345.86
3	551,250	110,250	200,000	356,650	277,875.05
4	578,812.50	115,762.50	200,000	370,982.50	261,535.89
5	607,753.13	121,550.63	200,000	386,031.63	246,247.21

Chap 8 - Prob. 15

Old unit: Depreciation costs $= \frac{600}{5} =$ \$120 per year

let $x =$ labor + maintenance costs

and $y =$ taxes + insurance costs

Manufacturing costs $= x + y$

New unit: Depreciation costs $= \frac{6000}{10} =$ \$600 per year

Manufacturing costs $= (x - 1000) + (y + 100)$

Savings with new unit $= (x + y) - \left[(x - 1000) + (y + 100) \right]$

$=$ \$900 per year

% replacement return $= \left[\frac{900 - (600 - 120)}{6000 - 600} \right] (100)$

$= 7.78\,\%$ Answer

Chap. 8 – Prob. 16

The project cost compounded over 10 years must be equal to the cash flow to the project compounded over 10 years using the given discounted-cash-flow rate of return

Batch system:

Future worth of project cost

$$S = 20{,}000\,(1.25)^{10} = \$186{,}200$$

Cash flow to project (based on end-of-year income)

$$S = 5600\left[\sum_{n=0}^{y-1}(1.25)^{y}\right] = (5600)(33.28) = \$186{,}400$$

This gives a good check on discounted-cash-flow rate of return

Continuous system

Future worth of project cost

$$S = 30{,}000\,(1.22)^{10} = (30{,}000)(7.3) = \$219{,}000$$

Cash flow to project (based on end-of-year income)

$$S = 7650\left[\sum_{n=0}^{y-1}(1.22)^{y}\right] = (7650)(27.72) = \$212{,}000$$

This gives a good check on discounted-cash-flow rate of return.

$$\text{Present worth} = \underbrace{\underbrace{\left[\frac{1}{(1+i)^n}\right]}_{\text{Discount factor}} \times \text{annual cash flow}}_{\text{Present value}} - \begin{array}{c}\text{initial}\\ \text{cost}\end{array}$$

For batch system

Year, n	Cash flow to project, $	Discount factor, i = 0.10	Present value, $
1	5600	0.909	
2	5600	0.827	
3	5600	0.751	
4	5600	0.683	
5	5600	0.621	
6	5600	0.564	
7	5600	0.514	
8	5600	0.467	
9	5600	0.424	
10	5600	0.386	
	Total	6.146	$34,400

Chap 8 - Prob. 16 (Cont'd)

Present value = (5600) (6.146) = \$34,400
Present worth = 34,400 − 20,000 = \$14,400

For continuous system :
Same discount factor applies
Present value = (7650) (6.146) = \$47,000
Present worth = 47,000 − 30,000 = \$17,000

The present worth values for both the batch and continuous systems check.

Rate of return :

A comparison of investments shows that an additional investment of \$10,000 yields incremental benefits of 7650 − 5600 or \$2050 in years 1 to 10.

This is a discounted-cash-flow rate of return of 16%.

$$S = 2050\left[\sum_{n=0}^{y-1}(1.16)^y\right] = 2050(21.4) = (10{,}000)(1.16)^{10} = \$44{,}000$$

Because the 16% incremental return exceeds the required return of 10%, the continuous system should be selected, as shown by present worth calculations.

Answer

Chap. 8 – Prob. 17

On the basis of the positive present worth at 10% the lease-and-water-flood arrangement should be undertaken. There is no rate of interest (less than infinity) which will yield a zero present worth. Therefore, if the board of directors understood only discounted-cash-flow rate of return, it would be necessary to reeducate the board to fully comprehend the limitations of the discounted-cash-flow rate-of-return method and demonstrate the effectiveness of the present-worth method before this proposal could be presented and fully evaluated.

Answer

Chap. 8 – Prob. 18

(a) A capital investment of $100 million is spent over three years. Assuming that the investment in each of the three years are equal to $33,333,333, we obtain the following depreciation every year:

Year	Amount depreciated from 1st year capital	Amount depreciated from 2nd year capital	Amount depreciated from 3rd year capital	Total
1	3,333,333	0	0	3,333,333
2	3,333,333	3,703,704	0	7,037,037
3	3,333,333	3,703,704	4,166,667	11,203,704
4	3,333,333	3,703,704	4,166,667	11,203,704
5	3,333,333	3,703,704	4,166,667	11,203,704
6	3,333,333	3,703,704	4,166,667	11,203,704
7	3,333,333	3,703,704	4,166,667	11,203,704
8	3,333,333	3,703,704	4,166,667	11,203,704
9	3,333,333	3,703,704	4,166,667	11,203,704
10	3,333,333	3,703,704	4,166,667	11,203,704
				Total = $100,000,000

Note that the investment of the 1st year is depreciated over 10 years, the investment of the second year is depreciated over 9 years and the investment of the third year is depreciated over 8 years.

Chap 8-Prob 18 (Cont'd)

Net profit = (sales − operating expenses − depreciation) (1 − tax rate)

Tax rate = 0.35

Year	Sales, \$	Expenses, \$	Depreciation, \$	Net Profit, \$
1	75×10^6	100×10^6	3,333,333	−18,416,666
2	112.5×10^6	100×10^6	7,037,037	3,550,926
3	150×10^6	100×10^6	11,203,704	25,217,592
4	150×10^6	100×10^6	11,203,704	25,217,592
5	150×10^6	100×10^6	11,203,704	25,217,592
6	150×10^6	100×10^6	11,203,704	25,217,592
7	150×10^6	100×10^6	11,203,704	25,217,592
8	150×10^6	100×10^6	11,203,704	25,217,592
9	150×10^6	100×10^6	11,203,704	25,217,592
10	150×10^6	100×10^6	11,203,704	25,217,592
			100×10^6	186,874,996

$$ROI = \frac{\sum_{j=1}^{N} N_j / 10}{\text{Total investment}} \times 100$$

$$= \frac{186,874,996 / 10}{100 \times 10^6 + 20 \times 10^6} \times 100$$

$$= 15.5\ \%$$

Answer

(b) $$PBP = \frac{\text{Fixed capital investment}}{\text{Average cash flow}}$$

Cash flow = Net profits + depreciation

$$\text{Average cash flow} = \frac{186,874,996 + 100,000,000}{10} = 28,687,500$$

Fixed capital investment = \$100,000,000

$$PBP = \frac{100,000,000}{28,687,500} = 3.5 \text{ years}$$

Chap 8 – Prob 19

We first compute the depreciation for each year

Year	Amount depreciated from 1^{st} year capital	Amount depreciated from 2^{nd} year capital	Amount depreciated from 3^{rd} year capital	Total
1	2×10^6	0	0	2×10^6
2	2×10^6	3.33×10^6	0	5.33×10^6
3	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
4	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
5	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
6	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
7	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
8	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
9	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6
10	2×10^6	3.33×10^6	6.25×10^6	11.58×10^6

$$\text{Present worth} = PWF_{c\&,j}\left[(s_j - c_j - d_j)(1-\phi) + rec_j + d_j\right]$$

From Table 7-5, $PWF_{c\&,j} = \left(\dfrac{e^r - 1}{r}\right) e^{-rj}$

Assume $r = 0.1$

Year, j	s_j	c_j	d_j	$\left(\frac{e^r-1}{r}\right)e^{-rj}$	Present Worth
1	75×10^6	100×10^6	2×10^6	0.951	–14,788,050
2	112.5×10^6	100×10^6	5.33×10^6	0.861	8,601,820
3	150×10^6	100×10^6	11.58×10^6	0.779	28,474,787
4	150×10^6	100×10^6	11.58×10^6	0.705	25,769,865
5	150×10^6	100×10^6	11.58×10^6	0.638	23,320,814
6	150×10^6	100×10^6	11.58×10^6	0.577	21,091,081
7	150×10^6	100×10^6	11.58×10^6	0.522	19,080,666
8	150×10^6	100×10^6	11.58×10^6	0.473	17,289,569
9	150×10^6	100×10^6	11.58×10^6	0.428	15,644,684
10	150×10^6	100×10^6	11.58×10^6	0.387	14,146,011

Total = \$158,631,247 $\approx$ \$$158.63 \times 10^6$

Chap 8 – Prob 19 (cont'd)

Present worth of all capital investments

Year, j	Capital investment, F_j	$PWF_{v,j} = \left(\frac{e^r-1}{r}\right)e^{-rj}$	$PWF_{v,j} \times F_j$
1	20×10^6	0.951	19.02×10^6
2	30×10^6	0.861	25.83×10^6
3	50×10^6	0.779	38.95×10^6
		Total	$\$83.80\times10^6$

Net present worth = Present worth of all cash flow − present worth of all capital investments

$$= 158.63\times10^6 - 83.80\times10^6$$

$$= \$74.83\times10^6$$

Answer

CHAPTER 9

PROB. 9-1

From prob. description,

Objective Function = Annual Cost Dependent Upon No. of Effects

Decision Variable(s) = No. of Effects $\triangleq X$

Investment Cost = $18k\$ + 15k\$(X-1)$

$\therefore$ ① = Fixed Annual Charge ex. Depr. = $[18k\$+15k\$(X-1)]0.15$

② = Annual Maint. Charges = $[18k\$+15k\$(X-1)]0.05$

③ = Annual Depr. Charges = $[18k\$+15k\$(X-1)]0.10$

$$①+②+③ = [18k\$+15k\$(X-1)](0.30) = 4500X - 900\ \$$$

$$\begin{Bmatrix}\text{Annual}\\ \text{Stm. Cost}\end{Bmatrix} = \frac{2\times10^5\ \text{kg}\ H_2O\ \text{evap.}}{\text{oper. Day}} * \frac{\text{kg Stm}}{0.85X\ \text{kg}\ H_2O\ \text{evap}} * \frac{0.0033\$}{\text{kg. Stm}} * \frac{300\ \text{oper. Da.}}{\text{y.}}$$

$$\Rightarrow \begin{Bmatrix}\text{Annual}\\ \text{Stm. Cost}\end{Bmatrix} = \frac{232941\$}{X}$$

$$\therefore \begin{Bmatrix}\text{Annual Cost}\\ \text{Dependent}\\ \text{Upon No. of Effects}\end{Bmatrix} = C_A = 4500X - 900 + \frac{232941}{X}$$

OPTIMUM NO. OF EFFECTS SATISFIES

$$\left.\frac{dC_A}{dX}\right|_{X_{OPT}} = 0 = 4500 - \frac{232941}{X_{OPT}^2}$$

$$\therefore X_{OPT} = 7.19 \Rightarrow \boxed{X_{OPT} = 7\ \text{EFFECTS}}$$

Problem 9.2 Optimum Insulation Thickness

MathCad File

Input Data

$k_{ins} := 5 \cdot 10^{-2} \quad \frac{W}{s \cdot m} \qquad h_o := 11.4 \quad \frac{W}{s \cdot m^2}$

$T_{stm} := 480 \quad K \qquad T_{surr} := 295 \quad K$

$\Delta H_v := 1.912 \cdot 10^6 \quad \frac{J}{kg}$ Heat of Vaporization of Sat.Stm. at 480 K

$L := 300 \quad m \qquad D_o := 0.273 \quad m$

$c_{stm} := 0.004 \quad \frac{dollars}{kg} \qquad \text{Install\_Cost}(I_t) := 180 \cdot I_t \quad \frac{dollars}{m}$

Annual Fixed Costs

$AFC(I_t) := 300 \cdot \text{Install\_Cost}(I_t) \cdot 0.20$

Annual Steam Cost

Use heat resistances to determine overall h.t. coefficient as function of I_t. Neglect resistance except for insulation and outer film. First express the areas involved:

$A_o(I_t) := \pi \cdot (D_o + 2 \cdot I_t) \cdot L \quad m^2 \qquad A_m(I_t) := .5 \cdot [D_o + (D_o + 2 \cdot I_t)] \quad m^2$

$$UA_o(I_t) := \frac{(h_o \cdot A_o(I_t) \cdot k_{ins} \cdot A_m(I_t))}{k_{ins} \cdot A_m(I_t) + I_t \cdot h_o \cdot A_o(I_t)} \quad \frac{W}{K}$$

from rearranging overall resistance from individual resistances.

$q(I_t) := UA_o(I_t) \cdot (T_{stm} - T_{surr}) \quad W$

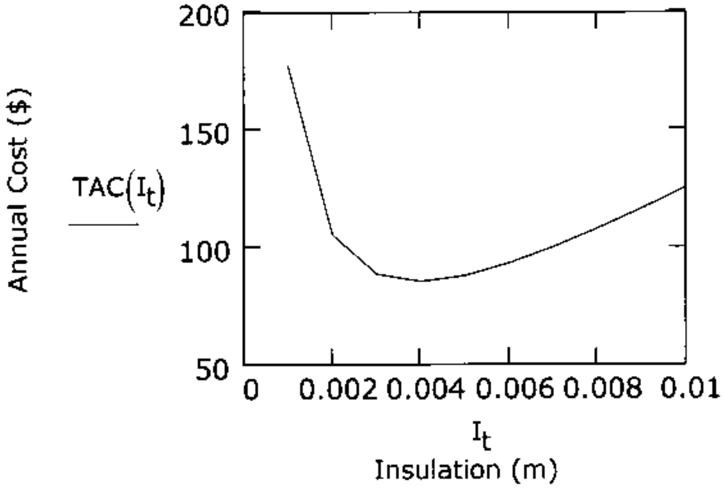
$$m_{stm}(I_t) := \frac{q(I_t)}{\Delta H_v} \quad \frac{kg}{s}$$

$ASC(I_t) := m_{stm}(I_t) \cdot c_{stm} \cdot 3600 \cdot 24 \cdot 365$ Annual Steam Cost

Total Annual Cost

$TAC(I_t) := AFC(I_t) + ASC(I_t) \quad \frac{dollars}{y}$

$I_t := 0.001, 0.002 .. 0.01$ Define a range of values for plotting



$j := 1, 2 .. 10$

$TABLE_{j,1} := 0.001 + (j - 1) \cdot 0.001$ $TABLE_{j,2} := TAC(TABLE_{j,1})$

I_t TAC

TABLE =

	1	2
1	$1 \cdot 10^{-3}$	177.242
2	$2 \cdot 10^{-3}$	105.318
3	$3 \cdot 10^{-3}$	88.459
4	$4 \cdot 10^{-3}$	85.413
5	$5 \cdot 10^{-3}$	87.9
6	$6 \cdot 10^{-3}$	93.156
7	$7 \cdot 10^{-3}$	99.995
8	$8 \cdot 10^{-3}$	107.824
9	$9 \cdot 10^{-3}$	116.312
10	0.01	125.263

Problem 9.3: Absorption Tower

Math Cad File

Input Data:

$F_V := 33 \ \frac{m^3}{s}$ $\quad M_W := 29.1 \ \frac{kg}{kgmol}$

$T := 400 \ K$ $\quad P := 111 \ kPa$

$HTU := 4.5 \ m$ $\quad NTU(G_s) := 1.05 \cdot G_s^{0.18}$ $\quad Op\_Hr := 8000 \ \frac{h}{y}$

$c_{tower} := 40 \ \frac{dollars}{m^3}$ $\quad f_{FAC} := 0.2$ factor for Fixed Annual Cost

$V\_OP\_Cost(G_s) := 1.22 \cdot 10^{-3} \cdot G_s^2 + 0.1098 \cdot G_s^{-1} + 0.0244 \cdot G_s^{-0.8} \quad \frac{dollars}{s}$

Solution:

$F_m := F_V \cdot \left(\frac{1}{22.4}\right) \cdot \left(\frac{P}{101.3}\right) \cdot \left(\frac{273}{T}\right) \cdot M_W$ $\quad F_m = 32.061 \ \frac{kg}{s}$

$G_s(D) := \frac{F_m}{0.25 \cdot \pi \cdot D^2}$

$z(D) := HTU \cdot NTU(G_s(D)) \ m$ Height of Packing

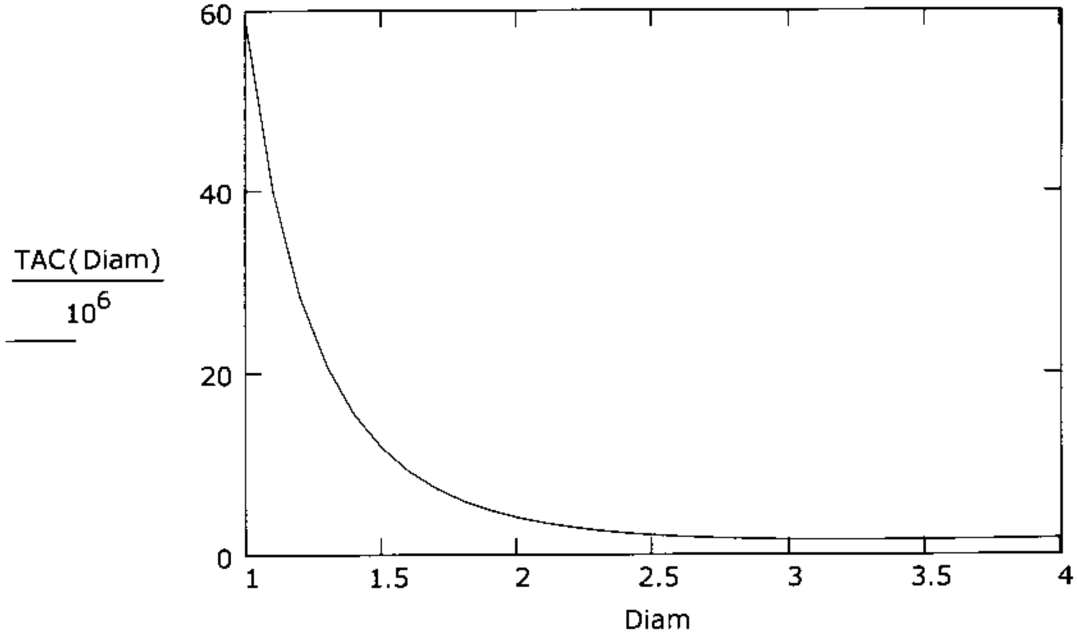
$Vol(D) := z(D) \cdot (0.25 \cdot \pi \cdot D^2) \ m^3$ Volume of Packing

$Ann\_Fixed\_Cost(D) := Vol(D) \cdot c_{tower} \cdot f_{FAC} \quad \frac{dollars}{y}$

$Ann\_Var\_Cost(D) := V\_OP\_Cost(G_s(D)) \cdot 3600 \cdot Op\_Hr \quad \frac{dollars}{y}$

$TAC(D) := Ann\_Fixed\_Cost(D) + Ann\_Var\_Cost(D)$

Diam := 1 , 1.1 .. 4



$F(D) := \frac{d}{dD} TAC(D)$ Take first derivative of Objective Function and set = 0

x := 2.5 Initial Guess Solve for Optimum Diameter below

Given

$F(x) = 0$

$D_{opt} := Find(x)$

$\boxed{D_{opt} = 3.272}$ m Optimum Diameter

$H_{opt} := z(D_{opt})$ $\boxed{H_{opt} = 6.012}$ m Optimum Height

$G_{sOPT} := G_s(D_{opt})$ $G_{sOPT} = 3.814$ $\frac{kg}{s \cdot m^2}$ Mass Velocity at Opt. Diam.

Costs at Optimum

$TAC(D_{opt}) = 1.581 \times 10^6$ \$/y Total Annual Cost

$Ann\_Fixed\_Cost(D_{opt}) = 404.342$ \$/y Total Annual Fixed Cost

$Ann\_Var\_Cost(D_{opt}) = 1.581 \times 10^6$ \$/y Total Annual Variable Cost

Optimum primarily depends upon variable costs in this problem.

PROB. 9.4

Let H_{op} = hrs. operation / yr.

δ_{ins} = insulation thickness (m)

c_{FC} = annual fixed cost associated w/ insulation / m of ins

c_H = cost of heat as \$/ J

ΔT = Temp. Driving Force (°K)

k = thermal conductivity (W/m-°K)

h_{air} = Air film heat transf. coefficient W/m²-°K

(a) Neglect air film resistance

∴ at steady state

$$q = \frac{Q}{A} = \frac{k\,\Delta T}{\delta_{ms}} \Rightarrow \frac{\text{Cost Heat}}{m^2\text{-y.}} = \frac{k\Delta T\, H_{op}\, c_H}{\delta_{ins}}$$

$$\frac{\text{Yearly Fixed Charges}}{m^2} = c_{FC}\,\delta_{ms} \quad \text{(By Prob. Statement)}$$

$$\Rightarrow \frac{\text{Total Annual Cost}}{m^2} = \frac{k\Delta T\, H_{op}\, c_H}{\delta_{ins}} + c_{FC}\,\delta_{ins}$$

$$\frac{d}{d\delta_{ms}}\left[\frac{\text{Total Ann. Cost}}{m^2}\right]\Bigg|_{\delta_{opt}} = 0 = -\frac{k\Delta T\, H_{op}\, c_H}{\delta_{opt}^2} + c_{FC}$$

$$\therefore \boxed{\delta_{opt} = \left[\frac{k\,\Delta T\, H_{op}\, c_H}{c_{FC}}\right]^{1/2}}$$

(b) with air film resistance included

$$q = U\Delta T \quad \text{where} \quad \frac{1}{U} = \frac{\delta_{ms}}{k} + \frac{1}{h_{AIR}}$$

$$\therefore U = \frac{k\, h_{AIR}}{k + \delta_{ins}\, h_{air}}$$

$$\Rightarrow q = \frac{k\, h_{air}}{k + \delta_{ms}\, h_{air}}\,\Delta T$$

PROB. 9.A (CON'T)

ⓑ CON'T

$$\sum_i \frac{\text{Tot. Ann. Cost}}{m^2} = \frac{k h_{air} \Delta T H_{op}}{k + \delta_{ins} h_{air}} c_H + \delta_{ins} c_{FC}$$

Taking the derivative with respect to δ_{ins}

$$\left.\frac{d(\text{Tot. Ann. Cost})}{d\,\delta_{ins}}\right|_{\delta_{OPT}} = 0 = -k h_{air} \Delta T c_H \frac{h_{air} H_{OP}}{(k + \delta_{OPT} h_{air})^2} + c_{FC}$$

$$\therefore \quad (k + \delta_{OPT} h_{air})^2 = \frac{k h_{air}^2 \Delta T c_H H_{OP}}{C_{FC}}$$

and $$\boxed{\delta_{OPT} = \left[\frac{k \Delta T c_H H_{OP}}{c_{FC}}\right]^{1/2} - \frac{k}{h_{air}}}$$

Problem 9.5 Evaporator

MathCad File

Objective Function = Production over 30 days
Decision Variable = Time onstream if cleaning time is 6 hours

Input Data

$R_1 := 5000 \ \frac{kg}{h}$ at $\theta_1 := 0$ and $R_2 := 2500 \ \frac{kg}{h}$ at $\theta_2 := 48$ h

$\theta_C := 6$ h

As noted on p.396, since the rate of production is proportional to U, $R = \frac{dP}{d\theta}$ will satisfy a relation of the form $\frac{1}{R^2} = a*\theta + b$

If we let $P(\theta_B)$ be the production up to θ_B, we can obtain the value by integrating the rate, R, from 0 to θ_B, namely,

$$P(\theta_B, a, b) := \int_0^{\theta_B} \frac{1}{(a \cdot \theta + b)^{0.5}} \, d\theta$$

Therefore

$$P(\theta_B, a, b) \rightarrow 2. \cdot \frac{(\theta_B \cdot a + b)^{\frac{1}{2}}}{a} - 2. \cdot \frac{b^{\frac{1}{2}}}{a}$$

Therefore

$$P_{30}(\theta_B, a, b, \theta_C) := P(\theta_B, a, b) \cdot \frac{(24 \cdot 30)}{\theta_B + \theta_C}$$

$$\frac{d}{d\theta_B} P_{30}(\theta_B, a, b, \theta_c) \rightarrow \frac{720.0}{(\theta_B \cdot a + b)^{\frac{1}{2}} \cdot (\theta_B + \theta_c)} - 720 \cdot \frac{\left[2. \cdot \frac{(\theta_B \cdot a + b)^{\frac{1}{2}}}{a} - 2. \cdot \frac{b^{\frac{1}{2}}}{a}\right]}{(\theta_B + \theta_c)^2}$$

The constants "a" and "b" may be evaluated from the given data

$$a := \frac{\left[\left(\frac{1}{R_2^{\,2}}\right) - \left(\frac{1}{R_1^{\,2}}\right)\right]}{\theta_2 - \theta_1} \qquad a = 2.5 \times 10^{-9}$$

$$b := \frac{1}{R_1^{\,2}} \qquad b = 40 \times 10^{-9}$$

Set the derivative P_{30} equal to zero and solve for the optimal θ_B

$$F(\theta_B) := \frac{d}{d\theta_B} P30(\theta_B, a, b, \theta_C)$$

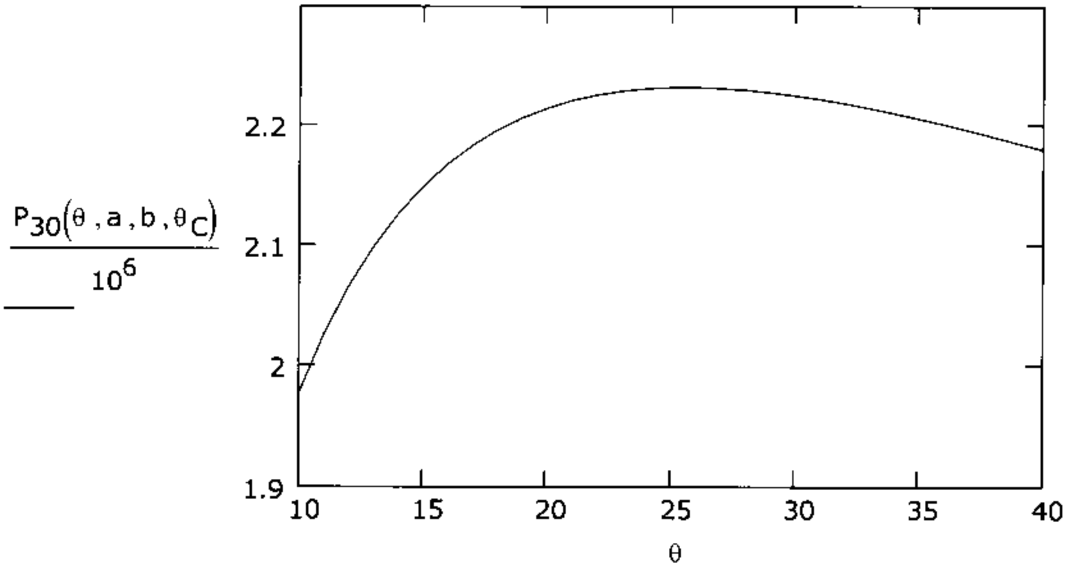
Initial Guess $\theta := 20$

Given

$F(\theta) = 0$

$\theta_{bOPT} := \text{Find}(\theta)$ $\boxed{\theta_{bOPT} = 25.596}$

$\theta := 10, 11 .. 40$ Define range for plotting



Problem 9.6 Solvent Extraction

MathCad File

Objective Function = Minimum Cost per Year
Decision Variable = Tower Diameter

Input Data:

$L_f := 40 \ \frac{m^3}{d}$ Feed Rate $\quad Op\_h := 24 \ \frac{h}{da} \quad Op\_d := 300 \ \frac{da}{y}$

$v_{allow} := 12.2 \ \frac{m}{h}$ recommended velocity of solvent and feed combined

Let F_{sf} = m^3 of solvent/m^3 of feed

$C_F(F_{sf}) := 8800 \cdot F_{sf}^2 - 51000 \cdot F_{sf} + 110000$ Fixed Charges ($/y)

$C_{OP}(F_{sf}) := 1.41 \cdot L_f \cdot F_{sf} \cdot Op\_d$ Operating Costs ($/y)

Solution

$C_{total}(F_{sf}) := C_F(F_{sf}) + C_{OP}(F_{sf})$ Total Cost ($/y)

The optimum value of F_{sf} is found by setting the derivative of the total cost equal to zero and solving for the optimum F_{sf}

$$g(F_{sf}) := \frac{d}{dF_{sf}} C_{total}(F_{sf}) \qquad \frac{d}{dF_{sf}} C_{total}(F_{sf}) \rightarrow 17600 \cdot F_{sf} - 34080.00$$

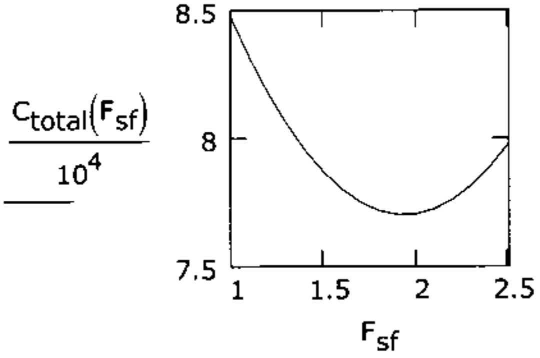
Setting this equal to zero, one obtains

$$F_{sfOPT} := \frac{34080}{17600} \qquad F_{sfOPT} = 1.936$$

Now we can determine the Tower diameter using the allowable total velocity

$$D_{opt} := \left[\left(\frac{4}{\pi}\right) \cdot \left(\frac{40}{24}\right) \cdot \left(1 + F_{sfOPT}\right)\right]^{0.5} \qquad \boxed{D_{opt} = 2.496} \ \text{m}$$

$F_{sf} := 1.0, 1.1 .. 2.5$



Problem 9.7 Optimum Pipe Diameter vs Flow Rate

MathCad File

For the range of viscosity considered, the viscosity term will range from 1 to 1.08 in the equation for diameters less than 0.0254m and from 1 to 1.077 in the equation for diameters greater than 0.0254 for the stated viscosity range. Therefore, we will plot $y = D_{I,opt}/\mu_{CP}{}^{0.025}$ and $z = D_{I,opt}/\mu_{CP}{}^{0.025}$ in the two cases where μ_{CP} is in cP.

From the equations on page 404 modified to account for viscosity in cP,

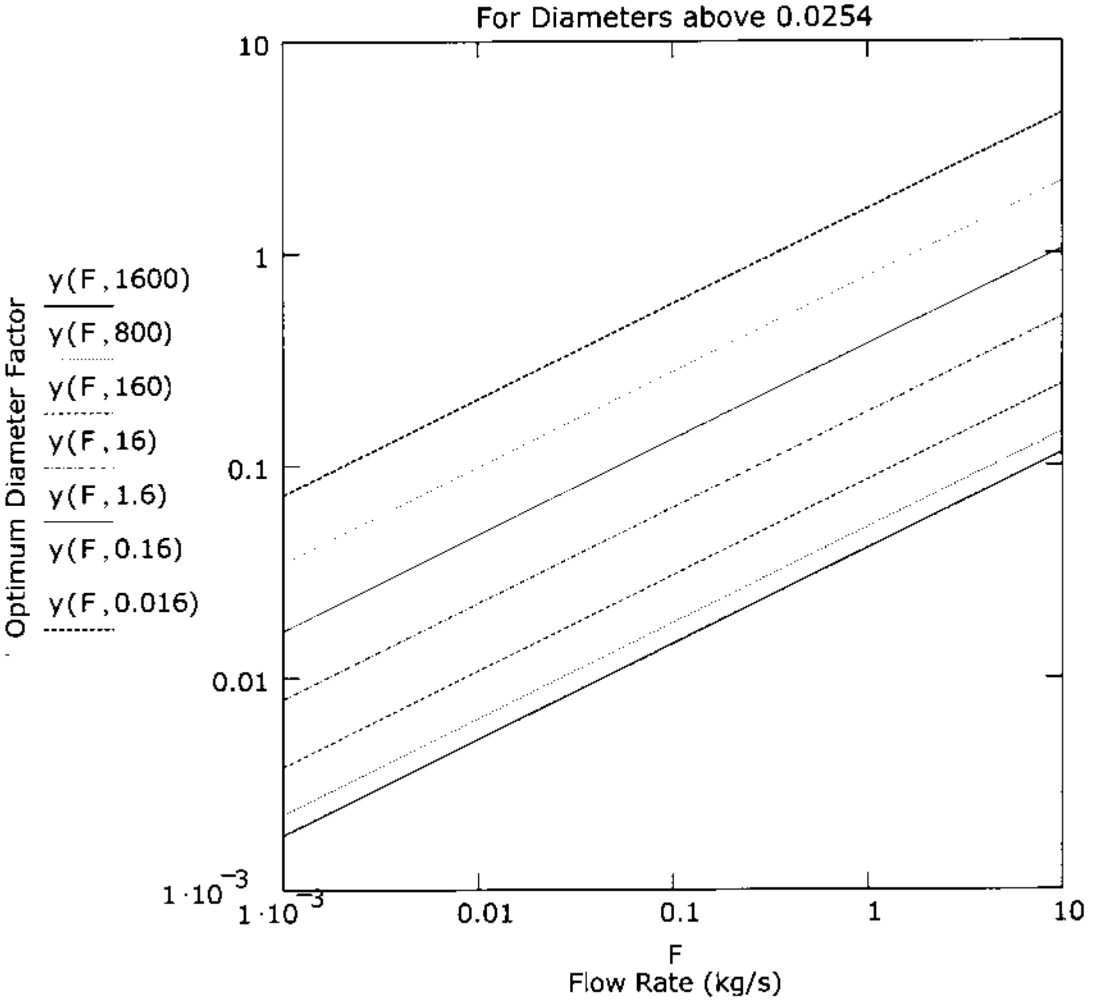
Turbulent Flow with Diameter greater than or equal to 0.0254

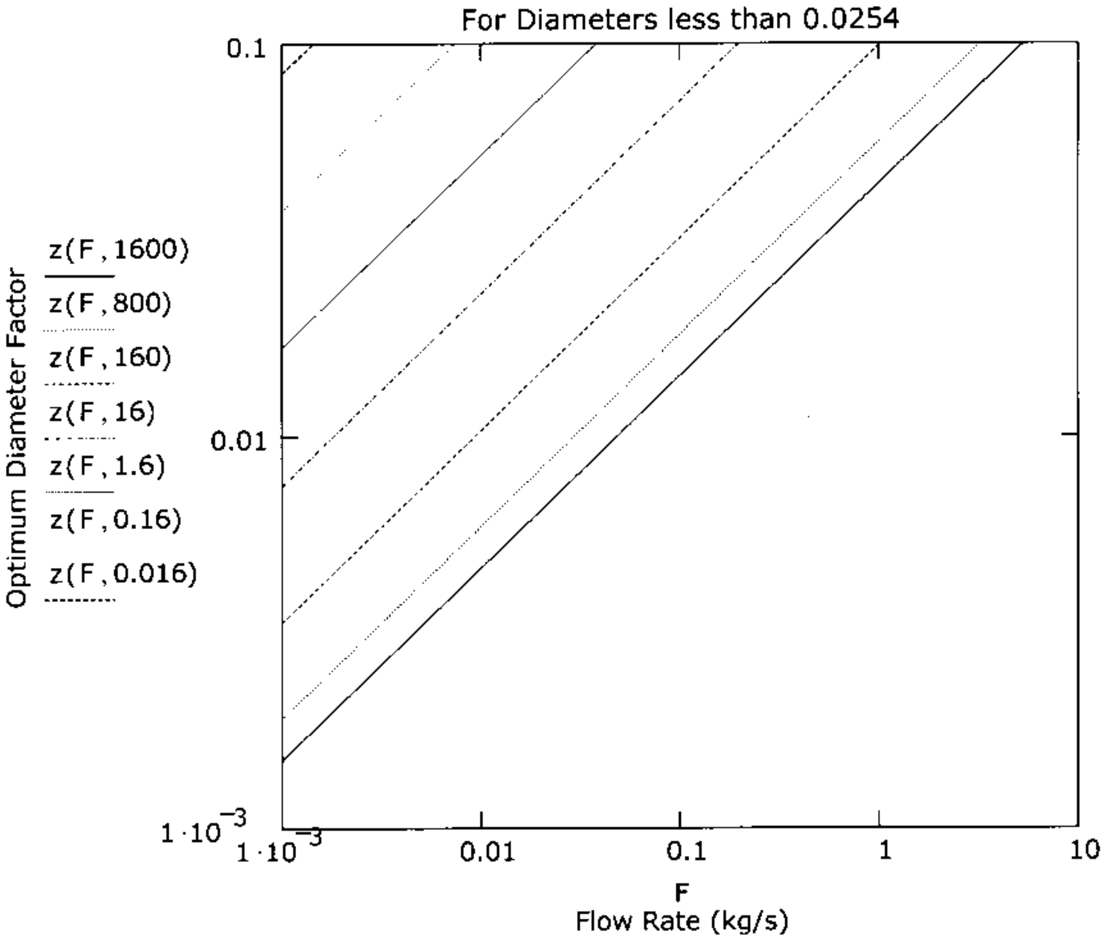
$$y(F_m, \rho) := 0.363 \cdot (1000)^{0.025} \cdot F_m{}^{0.45} \cdot \rho^{-0.32} \quad = D_{I,opt}/\mu_{CP}{}^{0.025}$$

Turbulent Flow with Diameter less than or equal to 0.0254

$$z(F_m, \rho) := 0.49 \cdot (1000)^{0.025} \cdot F_m{}^{0.49} \cdot \rho^{-0.35} \quad = D_{I,opt}/\mu_{CP}{}^{0.027}$$

$F := 0.001, 0.01 .. 10$ Define the range of flow rates to be plotted


For Diameters above 0.0254


For Diameters less than 0.0254

Problem 9.8 Optimum Pipe Diameter vs Velocity Math Cad File

The basis will be eqs. 9-76 and 9-77 on p.404, namely,

TURBULENT FLOW and diameter greater than or equal to 0.0254m

(1) $D_{I,opt}=0.363*(q_f^{0.45})*(\rho^{0.13})*(\mu^{0.025})$ with q_f in m^3/s

ρ in kg/m^3

μ in Pa-s

TURBULENT FLOW and diameter less than or equal to 0.0254m

(2) $D_{I,opt}=0.49*(q_f^{0.49})*(\rho^{0.14})*(\mu^{0.02})$ with q_f, ρ, μ in same units

If we substitute $q_f = v*\pi*(D_{I,opt})^2$ and $\mu = \mu_{CP}$(to place viscosity in units of cP) into the above equations, we obtain, respectively,

(1') $(D_{I,opt})^{0.1}=(7.221x10^{-2})*(v^{0.45})*(\rho^{0.13})*(\mu_{CP}^{0.025})$

(2') $(D_{I,opt})^{0.02}=(1.0347)*(v^{0.49})*(\rho^{0.14})*(\mu_{CP}^{0.027})$

Solving these equations for $D_{I,opt}$ and, similar to Problem 9.7, defining $y = D_{I,opt}/\mu_{CP}^{0.25}$ and $z = D_{I,opt}/\mu_{CP}^{1.35}$ to keep only ρ and v on the right hand side, we obtain:

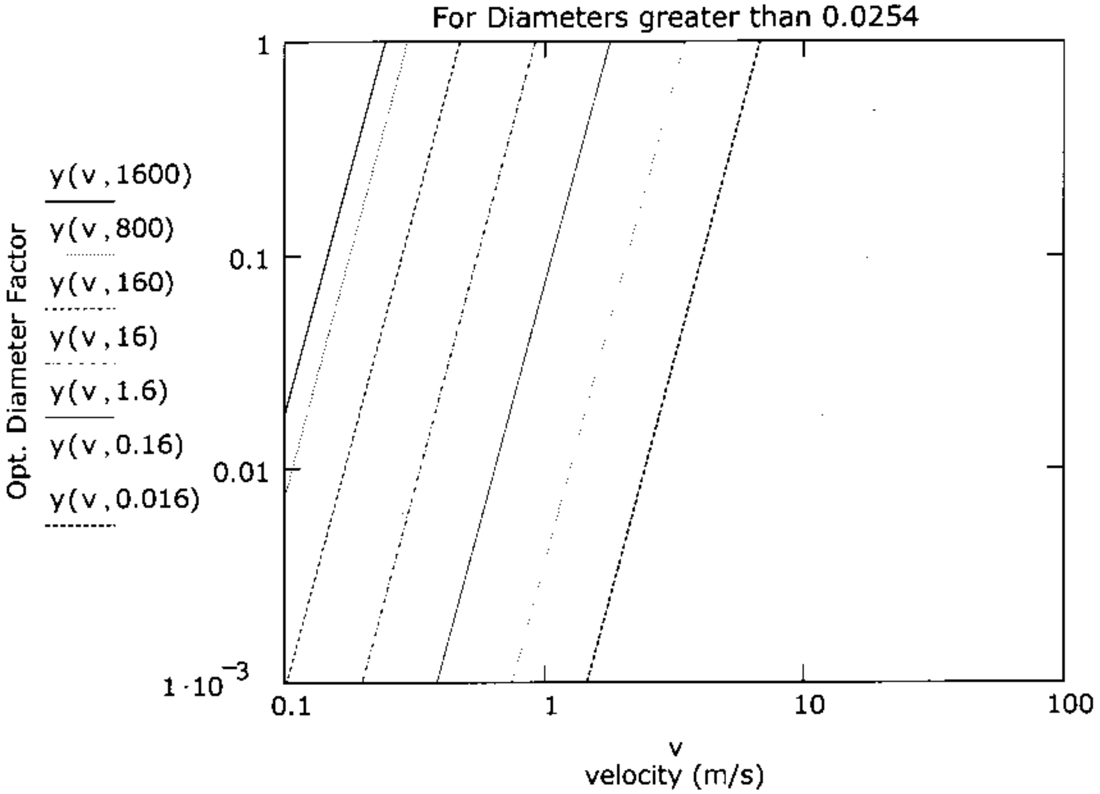
(1") $y = D_{I,opt}/\mu_{CP}^{0.25} = (3.855x10^{-2})*(v^{4.5})*(\rho^{1.3})$

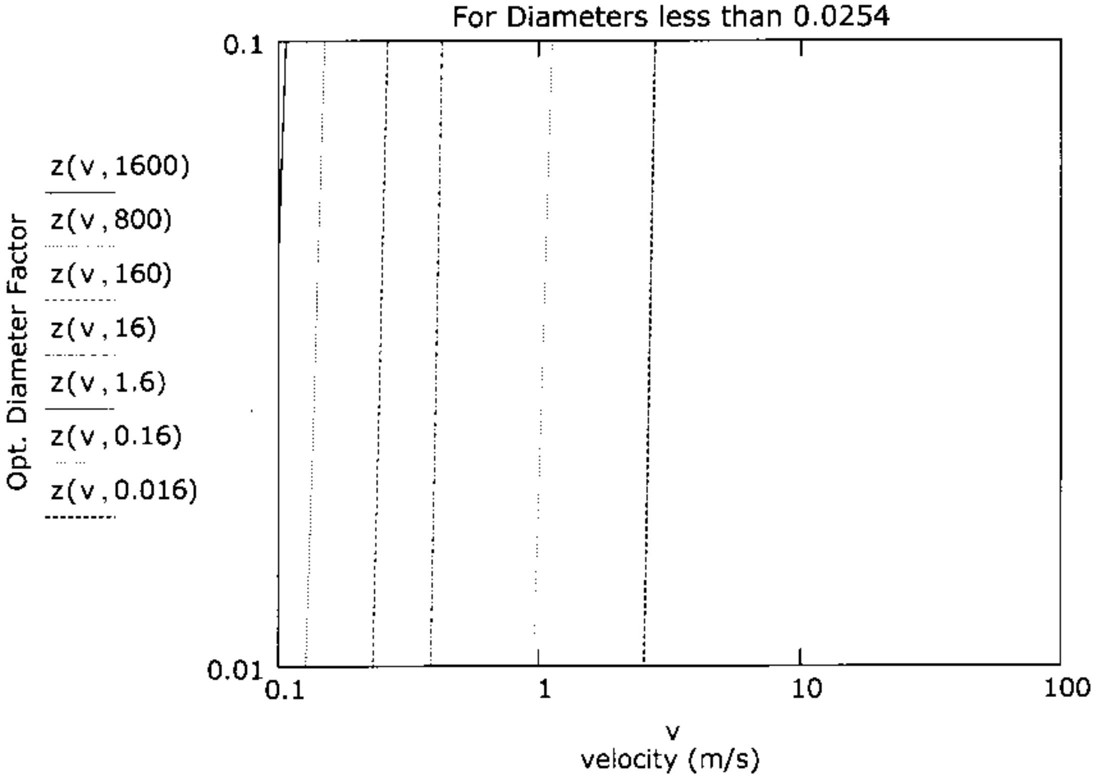
(2") $z = D_{I,opt}/\mu_{CP}^{1.35} = (5.505)*(v^{24.5})*(\rho^{7})$

$v := 0.1, 2 .. 30$ Define range of values of v for plotting

$y(v,\rho) := 3.885 \cdot 10^{-2} \cdot v^{4.5} \cdot \rho^{1.3}$ for diameters greater than 0.0254m

$z(v,\rho) := 5.505 \cdot v^{24.5} \cdot \rho^{7}$ for diameters less than 0.0254m


For Diameters greater than 0.0254
Opt. Diameter Factor
y(v, 1600)
y(v, 800)
y(v, 160)
y(v, 16)
y(v, 1.6)
y(v, 0.16)
y(v, 0.016)
1
0.1
0.01
1·10^−3
0.1
1
10
100
v
velocity (m/s)


For Diameters less than 0.0254
Opt. Diameter Factor
z(v, 1600)
z(v, 800)
z(v, 160)
z(v, 16)
z(v, 1.6)
z(v, 0.16)
z(v, 0.016)
0.1
0.01
0.1
1
10
100
v
velocity (m/s)

Problem 9.9 Evaporator - Minumum Total Cost MathCad File

Objective Function: Minimum Total Cost
Decision Variable: Cycle Time

This approach is discussed on pp.399-400.

The problem statement gives the correlation

$(1/U)^2=(6.88\text{x}10^{-5})*\theta_B + 0.186$ where U is in kJ/s-m^2-K and θ_B is in s.

If we want U in kJ/h-m^2-K and θ_B in hours, the correlation becomes:

$(1/U)^2=(6.88\text{x}10^{-5}/3600)*\theta_B + (0.186/3600^2)$

Therefore

$a := \frac{\left(6.88 \cdot 10^{-5}\right)}{(3600)}$ $\quad a = 1.911 \times 10^{-8}$ $\quad c := \frac{0.186}{(3600)^2}$ $\quad c = 1.435 \times 10^{-8}$

Other Input Data:

$\lambda := 2300 \ \frac{kJ}{kg}$ $\quad A := 37 \ m^2$ $\quad \Delta T := 40 \ C$

$\theta_C := 4 \ h$ $\quad C_C := 100$ \$ per cycle $\quad S_B := 20$ \$ per hour

Part a

$F_m := 30000 \ \frac{kgH2Oevaporated}{da}$

From equation 9-61

$$\theta_{Bopt} := \left(\frac{C_C}{S_B}\right) + \left[\left(\frac{2}{a \cdot S_B}\right) \cdot \left(a \cdot c \cdot C_C \cdot S_B\right)^{0.5}\right] \qquad \boxed{\theta_{Bopt} = 8.875} \ h$$

$$\theta_T := \theta_{Bopt} + \theta_C \qquad \boxed{\theta_T = 12.875} \ h$$

Check using available time

$H_{avail} := 30 \cdot 24 \ h$ $\qquad Q_H := F_m \cdot 30 \cdot \lambda \ \ kJ$

$$\theta_t := \left[\frac{\left(2 \cdot A \cdot H_{avail} \cdot \Delta T\right)}{\left(a \cdot Q_H\right)}\right] \cdot \left[\left(a \cdot \theta_{Bopt} + c\right)^{0.5} - c^{0.5}\right] \qquad \boxed{\theta_t = 16.653} \ h$$

Since the min. cost opt. is less than the available, we may use θ_{Bopt}

Part b

$F_{m2} := 37000 \ \frac{kgH2Oevaporated}{da}$

Using the results of part (a), we have for the available cycle time

$\theta_{t2} := \theta_t \cdot \left(\frac{F_m}{F_{m2}}\right)$ $\qquad \theta_{t2} = 13.503$ h

Therefore, the available cycle time is still greater that the cycle time for min. cost.

* If, however, the new rate had been 40000

$F_{m2} := 40000 \ \frac{kgH2Oevaporated}{da}$ $\qquad \theta_{t2} := \theta_t \cdot \left(\frac{F_m}{F_{m2}}\right)$ $\qquad \theta_{t2} = 12.49$ h

Therefore, the available cycle time would be less than the cycle time for min. cost and one would have to determine the cycle time by solving the following equation:

$\theta_b := 8$ Initial Guess

Given

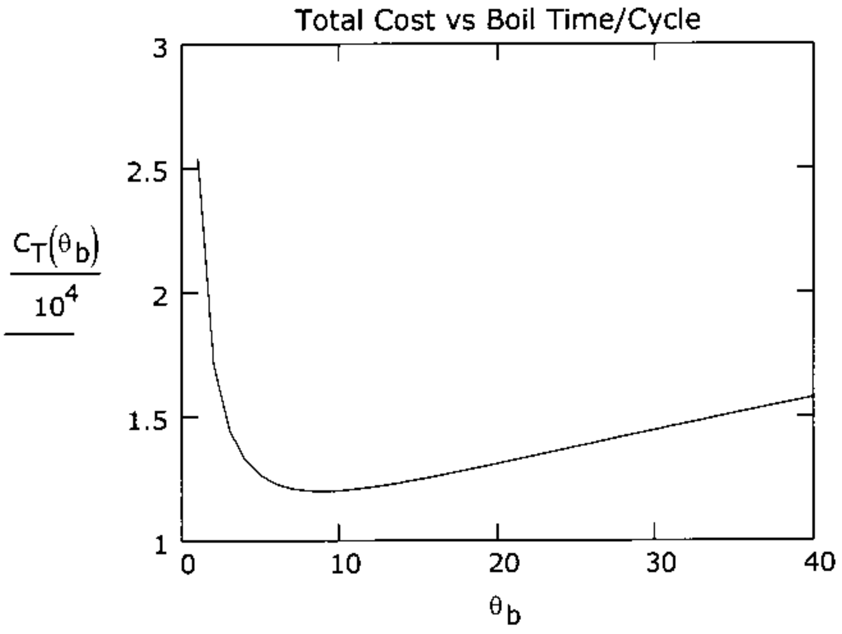
$$\theta_C + \theta_b = \left[\frac{(2 \cdot A \cdot H_{avail} \cdot \Delta T)}{(a \cdot Q_H)}\right] \cdot \left[(a \cdot \theta_b + c)^{0.5} - c^{0.5}\right]$$

$\theta_b := \mathrm{Find}(\theta_b)$ $\qquad \boxed{\theta_b = 2.083}$ h

If the initial guess had been 14, the solution would have converged to 32.5 h

$$C_T(\theta_b) := \frac{\left[a \cdot Q_H \cdot (C_C + S_B \cdot \theta_b)\right]}{2 \cdot A \cdot \Delta T \cdot \left[(a \cdot \theta_b + c)^{0.5} - c^{0.5}\right]}$$

$\theta_b := 1, 2 .. 40$ range for plotting



Problem 9.10 Maximum Profit per kg Product Math Cad File

Problem Info:

C_X := 0.2 $/kg X C_Y := 0.1 $/kg Y SP_Z := 1.75 $/kg Z

C_Other := 0.5 • SP_Z C_Other = 0.875

If we set $kg_Z = 1$ in the given relationship between kg_X, kg_Y, and kg_Z, we can solve for kgX_Z and kgY_Z which we define as the kg_X/kg_Z and kg_Y/kg_Z respectively. The given relationship, therefore becomes:

$$1 = (1.5)^2[1.1*kgX\_Z + 1.3*kgY\_Z - kgX\_Z*kgY\_Z]$$

Solving for kgY_Z as a funtion of kgX_Z, we obtain:

$$kgY\_Z = [1 - 2.475*kgX\_AZ] / [2.925 - 2.25*kgX\_Z]$$

Defining this relation as a function

$$kgY\_Z(kgX\_Z) := \frac{(1 - 2.475 \cdot kgX\_Z)}{(2.925 - 2.25 \cdot kgX\_Z)}$$ kgY_Z as a function of kgX_Z

C_kgZ(kgX_Z) := C_X • kgX_Z + C_Y • kgY_Z(kgX_Z) + C_Other *Cost/kg Z*

P_kgZ(kgX_Z) := SP_Z − C_kgZ(kgX_Z) Profit/kg Z

Note: In this case maximizing the profit is the same as minimizing the cost. Also, the optimum is not a function of the selling price or the cost of items other than the raw material costs. Therefore, the optimum coulod be determined by minimizing the raw material costs.

To determine the optimum value of kgX_Z, we take the derivative of the Profit function (or, if one wishes the raw material cost function), set it equal to zero, and solve for kgX_Z

$$\frac{d}{dkgX\_Z}P\_kgZ(kgX\_Z) \rightarrow -.2 + \frac{.2475}{(2.925 - 2.25 \cdot kgX\_Z)} - .225 \cdot \frac{(1 - 2.475 \cdot kgX\_Z)}{(2.925 - 2.25 \cdot kgX\_Z)^2}$$

Defining the optimum value of kgX_Z as x_{opt}

$$F(kgX\_Z) := \frac{d}{dkgX\_Z}P\_kgZ(kgX\_Z)$$ Define the function corresponding to the derivative

Set up to solve for x_{opt}

$x_{opt} := 1.6$ Initial Guess

Given

$F(x_{opt}) = 0$

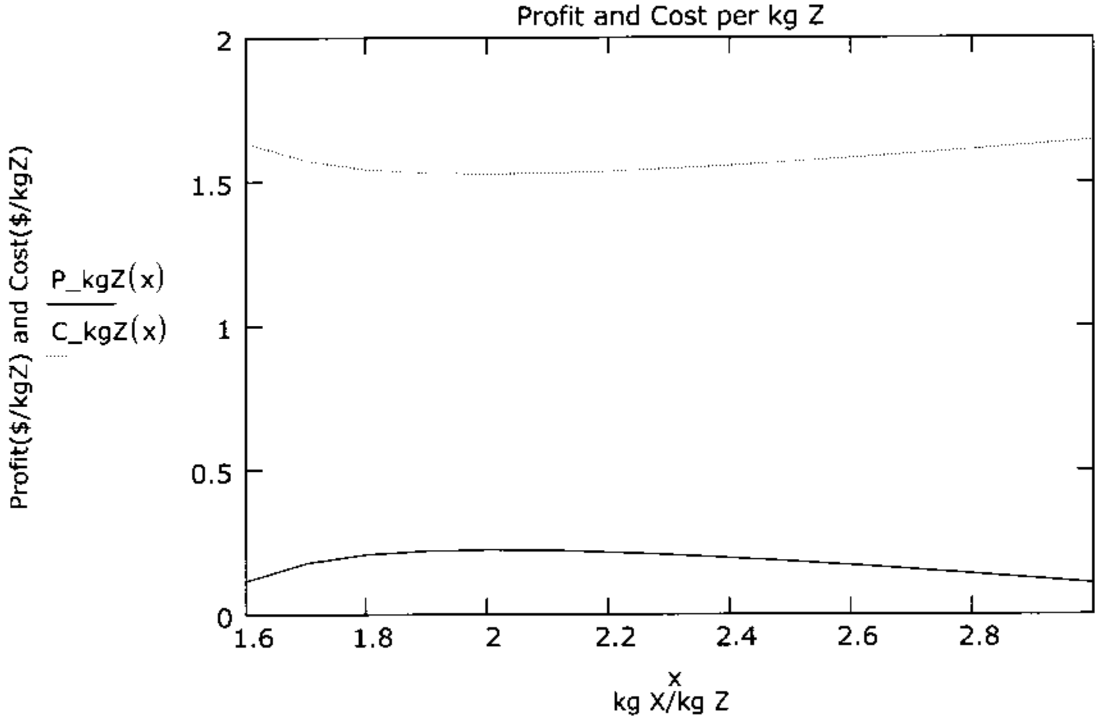
$x_{opt} := Find(x_{opt})$ $\boxed{x_{opt} = 2.002}$ Optimum kg_X/kg_Z

$y_{opt} := kgY\_Z(x_{opt})$ $\boxed{y_{opt} = 2.504}$ Optimum kg_Y/kg_Z

$\boxed{C\_kgZ(x_{opt}) = 1.526}$ Minimum Total $Cost/kg_Z$

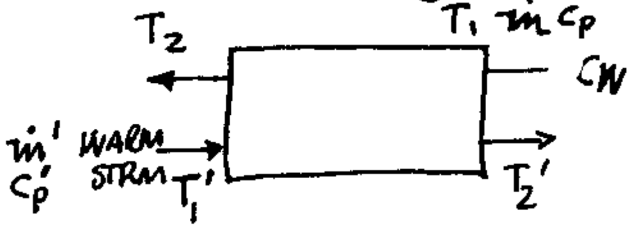
$\boxed{P\_kgZ(x_{opt}) = 0.224}$ Maximum $Profit/kg_Z$

$x := 1.6, 1.7 .. 3$ Range for plotting



PROB 9-11

Use nomenclature of p. 407 with $F_G \triangleq (\Delta T_{ACTUAL}/\Delta T_{lm})$ and primes to designate items from stream being cooled



$$\dot{q} = \dot{m} c_p (T_2 - T_1)$$
$$= \dot{m}' c_p' (T_1' - T_2')$$
$$= U A F_G \Delta T_{lm} \qquad \text{where } \Delta T_{lm} \triangleq \frac{(T_1' - T_2) - (T_2' - T_1)}{\ln\left[\frac{T_1' - T_2}{T_2' - T_1}\right]}$$

$$\begin{Bmatrix}\text{Total Annual}\\ \text{Cost}\end{Bmatrix} = C_T = \begin{Bmatrix}\text{Exchr. Fixed}\\ \text{Cost}\end{Bmatrix} + \begin{Bmatrix}\text{Cooling Water}\\ \text{Cost (note in text)}\end{Bmatrix}$$

But $$\begin{Bmatrix}\text{Exchr. Fixed}\\ \text{Cost}\end{Bmatrix} = K_F C_A A = K_F C_A \frac{\dot{q}}{U F_G \Delta T_{lm}}$$

$$\begin{Bmatrix}\text{Cooling Water}\\ \text{Cost}\end{Bmatrix} = \dot{m} H_y C_W = H_y C_W \frac{\dot{q}}{c_p (T_2 - T_1)}$$

$$\therefore C_T = \dot{q} \left[\frac{H_y C_W}{c_p (T_2 - T_1)} + \frac{K_F C_A}{U F_G \Delta T_{lm}} \right] \qquad ①$$

To find the T_2 to minimize C_T, say T_{2opt}, we can solve the equation

$$\left.\frac{d C_T}{d T_2}\right|_{T_{2opt}} = 0 \qquad ②$$

From ① and ②

$$\left.\frac{dC_T}{dT_2}\right|_{T_{2opt}} = \dot{q} \left[\frac{H_y C_W}{c_p} \left\{ \frac{-1}{(T_{2_{opt}} - T_1)^2} \right\} + \frac{K_F C_A}{U F_G} \left\{ \frac{-1}{\Delta T_{lm}^2} \frac{d}{dT_2} (\Delta T_{lm}) \right\} \right] = 0$$

PROB. 9-11 (con't)

$$\therefore -\frac{d}{dT_2}(\Delta T_{lm})\Big|_{T_{2,opt}} = \frac{H_y C_W U F_G}{c_p K_F C_A}\,\frac{(\Delta T_{lm})^2}{(T_{2opt}-T_1)^2} \qquad ③$$

If we let $\Delta T_h = T_1' - T_2$ & $\Delta T_c = T_2' - T_1$
and then $u = \Delta T_h - \Delta T_c$ & $v = \ln\frac{\Delta T_h}{\Delta T_c} = \ln \Delta T_h - \ln \Delta T_c$

$$\frac{d}{dT_2}(\Delta T_{lm}) = \frac{d}{dT_2}\left(\frac{u}{v}\right) = \frac{1}{v}\,\frac{du}{dT_2} - \frac{u}{v^2}\,\frac{dv}{dT_2}$$

But $\frac{du}{dT_2} = -1$ & $\frac{dv}{dT_2} = \frac{-1}{\Delta T_h}$

and substituting into ③ with $\Delta T_{lm} = \frac{\Delta T_h - \Delta T_c}{\ln[\Delta T_h/\Delta T_c]}$

$$\frac{1}{\ln\left[\frac{\Delta T_h}{\Delta T_c}\right]} + \frac{1}{\left\{\ln\left[\frac{\Delta T_h}{\Delta T_c}\right]\right\}^2}\,\frac{1}{\Delta T_h} = \frac{H_y C_W U F_G}{c_p K_F C_A}\,\frac{1}{(T_{2opt}-T_1)^2}\left\{\frac{\Delta T_h - \Delta T_c}{\ln\left[\frac{\Delta T_h}{\Delta T_c}\right]}\right\}^2$$

Rearranging and cancelling like terms, we obtain

$$\frac{H_y C_W U F_G}{c_p K_F C_A} = (T_{2opt}-T_1)^2\left\{\frac{\ln(\Delta T_{hopt}/\Delta T_c)}{(\Delta T_{hopt}-\Delta T_c)^2} + \frac{1}{\Delta T_{hopt}(\Delta T_{hopt}-\Delta T_c)^2}\right\}$$

$$= \frac{(T_{2opt}-T_1)^2}{(\Delta T_{hopt}-\Delta T_c)^2}\left\{\frac{(\Delta T_{hopt}-\Delta T_c)}{\Delta T_{lm,opt}} + \frac{1}{\Delta T_{opt}}\right\}$$

where $\Delta T_{hopt} = T_1' - T_{2opt}$ & $\Delta T_{lmopt} = \frac{\Delta T_{hopt} - \Delta T_c}{\ln\left[\frac{\Delta T_{hopt}}{\Delta T_c}\right]}$

PROB. 9-12

Temp. of Sat. Stm at 800 kPa ≅ $170.4^\circ C = 443.6\,^\circ K$

$k \cong 0.066 \frac{W}{m\text{-}K}$ (estimated from C. Geankoplis "Transport Proc. and Unit Ops., 3ed., Prentice-Hall) at 148°C

$H_y = 365 * 24 * 3600 = \quad 3.154 \times 10^7\ s/y$

$T_{surr} = 300\,^\circ K$

$\therefore\ \Delta T = 443.6 - 300 = 143.6\,^\circ K$

$C_{ins} = 700\ \$/m^3$

$C_{FIXED} = 0.2 * [\text{Initial Investment}]$

$C_{stm} = 1.42 \times 10^{-9}\ \$/J \implies b_c = 1.42 \times 10^{-6}\ \$/kJ$

$a_c = (0.2)(700) = 140\ \$/m^3\text{-}y$

$$\therefore \left(\frac{k\, b_c\, H_y \Delta T}{a_c}\right)^{1/2} = \left[\frac{(0.066 \times 10^{-3})(1.42 \times 10^{-6})(3.154 \times 10^7)(143.6)}{140}\right]^{1/2}$$

$$= 0.055 = \text{FACTOR}$$

FROM TABLE

	0.025 m	0.051 m
@ FACTOR = 0.03	0.010	0.013
@ FACTOR = 0.06	0.024	0.028

BY LINEAR INTERPOLATION

Opt. Econ. Thickness of Insulation = 0.0236

∴ SELECT 0.025 m insulation available

Problem 9.13 Optimum Cycle Time for Maximum Profit MathCad File

Input Information:

$F := 70\ \frac{kg}{da}$ $Op\_da := 300\ \frac{Oper\_da}{y}$ $c_{feed} := 5.50$ \$/kg

$c_{oper} := 300$ \$/oper.da *Cost for Operations* $c_{regen} := 800$ \$/regen *Cost of Regeneration*

$A\_C\_Other := 100000$ \$/y Annual Fixed, O/H, etc.

$Y(\theta) := 0.87 \cdot \theta^{-0.25}\ \frac{kgProd}{kgFeed}$ Product Yield

$P_S := 31$ \$/kgProd *Product Value*

$Prod\_Rate(\theta) := F \cdot Y(\theta)$ *Production Rate*

Assuming the yield for the first day is 0.87

$$Prod(\theta_D) := \int_1^{\theta_D} Prod\_Rate(\theta)\, d\theta + F \cdot Y(1)$$

$$Prod(\theta_D) \rightarrow 81.20 \cdot \theta_D^{\frac{3}{4}} - 20.30$$

$A\_Inc(\theta_D) := Prod(\theta_D) \cdot \frac{Op\_da}{\theta_D} \cdot P_S$ Annual Income as Function of θ_D

$A\_C\_Feed := F \cdot c_{feed} \cdot Op\_da$ Annual Cost of Feed

$A\_C\_Oper := c_{oper} \cdot Op\_da$ Annual Cost for Operations

$A\_C\_Regen(\theta_D) := c_{regen} \cdot \frac{Op\_da}{\theta_D}$ Annual Cost for Regenerations

$A\_C\_Total(\theta_D) := A\_C\_Feed + A\_C\_Oper + A\_C\_Regen(\theta_D) + A\_C\_Other$

$Ann\_Profit(\theta_D) := A\_Inc(\theta_D) - A\_C\_Total(\theta_D)$

The optimum θ_D is found by setting the first derivative of the annual profit to zero and solving for $\theta_{D_{opt}}$

$$F(\theta_D) := \frac{d}{d\theta_D} Ann\_Profit(\theta_D)$$

$$\frac{d}{d\theta_D} Ann\_Profit(\theta_D) \rightarrow \frac{5.664 \cdot 10^5}{\theta_D^{\frac{5}{4}}} - 9300 \cdot \frac{\left(81.20 \cdot \theta_D^{\frac{3}{4}} - 20.30\right)}{\theta_D^2} + \frac{240000}{\theta_D^2}$$

Solve for θ_{Dopt}

Initial Guess $\theta_D := 4$

Given

$F(\theta_D) = 0$

$\theta_{Dopt} := \text{Find}(\theta_D)$ $\boxed{\theta_{Dopt} = 2.986}$ da

$\text{Ann\_Profit}(\theta_{Dopt}) = 1.254 \times 10^5$

$\text{A\_Inc}(\theta_{Dopt}) = 5.113 \times 10^5$

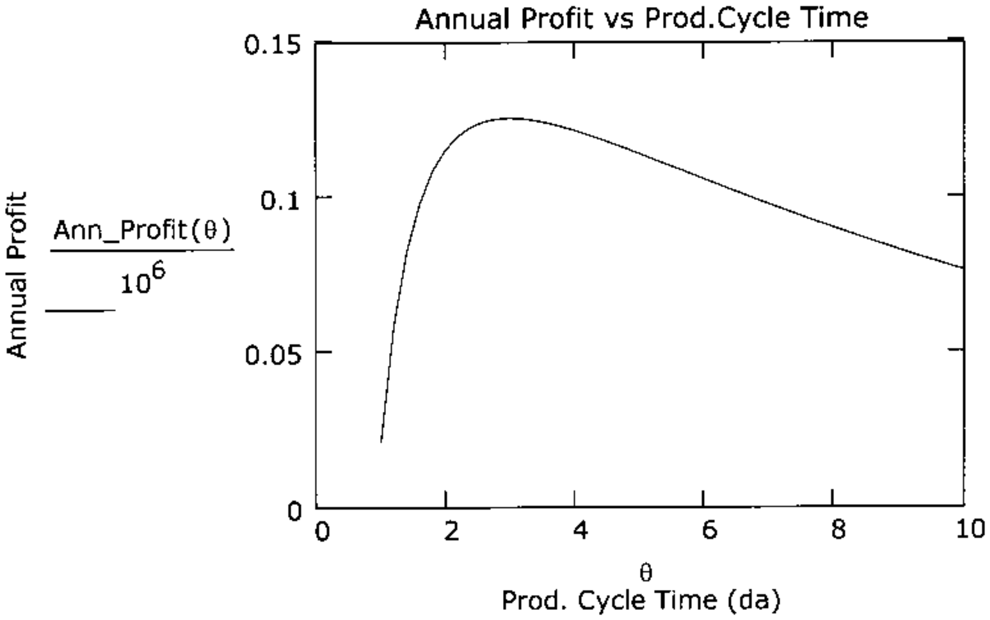
$\text{A\_C\_Feed} = 1.155 \times 10^5$

$\text{A\_C\_Regen}(\theta_{Dopt}) = 8.039 \times 10^4$

$\text{A\_C\_Oper} = 9 \times 10^4$

$\text{A\_C\_Total}(\theta_{Dopt}) = 3.859 \times 10^5$

$\theta := 1, 1.2 .. 10$ range for plotting



PROB. 9-14

$$\dot{Q} = \frac{\Delta T_{ov}}{R_{th,ov}}$$

$$R_{th,ins} = \frac{\ln(r_2/r_1)}{2\pi L k_{in}}$$ = thermal resist. of insulation

$$R_{th,out} = \frac{1}{(h_c+h_r)_c 2\pi L r_2}$$ = thermal resist. due to conv. and radiation from outer portion of insulation

$$\therefore R_{th,ov} = \left[\frac{\ln(r_2/r_1)}{k_{in}} + \frac{1}{(h_c+h_r)_c r_2}\right]\frac{1}{2\pi L}$$

$$\Rightarrow \dot{Q} = \frac{2\pi L\, \Delta T_{ov}}{\left[\frac{\ln(r_2/r_1)}{k_{in}} + \frac{1}{(h_c+h_r)_c r_2}\right]}$$

To find r_{2opt}

$$\left.\frac{d\dot{Q}}{dr_2}\right|_{r_{2opt}} = 0$$

But

$$\frac{d\dot{Q}}{dr_2} = 2\pi L \Delta T_{ov}(-1)\frac{1}{\left[\frac{\ln(r_2/r_1)}{k_{in}} + \frac{1}{(h_c+h_r)_c r_2}\right]^2}\left\{\qquad\qquad\right\}$$

$$\boxed{\frac{1}{r_2}\frac{1}{k_{in}} + \frac{(-1)}{(h_c+h_r)_c}\frac{1}{r_2^2}}$$

Setting this derivative to zero, we obtain

$$\frac{1}{k_{in} r_{2opt}} = \frac{1}{(h_c+h_r)_c r_{2opt}^2}$$

$$\therefore r_{2opt} = \frac{k_{in}}{(h_c+h_r)_c} \quad ; \quad \boxed{D_{opt} = 2r_{2opt} = \frac{2k_{in}}{(h_c+h_r)_c}}$$

PROB 9-15

Using the nomenclature of pp. 401–404, from eq. 9-66 for turbulent flow,

$$C_{pumping} = \frac{1.248\times10^{-4}\, q_f^{2.84}\, \rho^{0.84}\, \mu_c^{0.16}\, K(1+J)H_y}{D_i^{4.84}\, E} + B'$$

From eq. 9-69, if K_F is factor for fixed charges (incl. maintenance) expressed as fraction of initial cost of installed pipe,

$$C_{pipe} = (1+F)X\left(\frac{D_i}{0.0254}\right)^{n} K_F$$

In order to use the same variables as text in eq. 9-80, we substitute

$$q_f = \frac{\dot{m}}{\rho} \Rightarrow q_f^{2.84}\,\rho^{0.84} \text{ becomes } \frac{\dot{m}^{2.84}}{\rho^{2}}$$

$$J = L'_e D_i$$

$$H_y = 24Y \qquad X = X'$$

$$\therefore C_{pumping} = \frac{2.995\times10^{-3}\, \dot{m}^{2.84}\, \mu_c^{0.16}\, K Y(1+L'_e D_i)}{\rho^{2}\, E\, D_i^{4.84}} + B'$$

Using the factor M we may express the initial cost of the pumping installation as:

$$C_{pump.install} = M\, C_{pumping}$$

$$\therefore \begin{Bmatrix}\text{Total Annual}\\ \text{Cost}\end{Bmatrix} = C_T = C_{pumping}(1-\Phi) \qquad \boxed{\text{POWER COST}}$$

$$+\left[C_{pipe}(a+b) + MC_{pumping}(a'+b')\right](1-\Phi) \qquad \boxed{\text{Depr + Maint. Cost}}$$

$$+\left[C_{pipe} + MC_{pumping}\right]Z \qquad \boxed{\text{Cost of Capital}}$$

PROB 9-15 (con't.)

or, upon rearranging

$$C_T = C_{pipe}\underbrace{[(a+b)(1-\Phi)+Z]}_{A_1} + C_{pumping}\underbrace{[(1-\Phi)+M(a'+b')(1-\Phi)+MZ]}_{A_2}$$

or $C_T = A_1 C_{pipe} + A_2 C_{pumping}$

The optimal pipe Diameter is determined from the equation

$$\left.\frac{dC_T}{dD_i}\right|_{D_{opt}} = A_1 \left.\frac{dC_{pipe}}{dD_i}\right|_{D_{opt}} + A_2 \left.\frac{dC_{pumping}}{dD_i}\right|_{D_{opt}} = 0 \quad ①$$

From the expressions for C_{pipe} and $C_{pumping}$

$$\left.\frac{dC_{pipe}}{dD_i}\right|_{D_{opt}} = n(1+F)X'(D_{opt})^{n-1}\frac{1}{(0.0254)^n}$$

$$\left.\frac{dC_{pumping}}{dD_i}\right|_{D_{opt}} = -\frac{2.995\times10^{-3}\,\dot{m}^{2.84}\mu_c^{0.16}KY}{\rho^2 E}\left[3.84D_{opt}^{-4.84}L_e' + 4.84D_{opt}^{-5.84}\right]$$

$$= -\frac{1.45\times10^{-2}\,\dot{m}^{2.84}\mu_c^{0.16}KY}{\rho^2 E}\left[D_{opt}^{-5.84} + 0.7934L_e' D_{opt}^{-4.84}\right]$$

Substituting these expressions for the derivatives into ① above and multiplying through by $D_{opt}^{5.84}$, we obtain

$$\frac{D_{opt}^{4.84+n}}{1+0.7934L_e' D_{opt}} = \frac{(1.45\times10^{-2})(0.0254)^n\,\dot{m}^{2.84}\mu_c KY}{\rho^2 E\, n(1+F)X'}\ \frac{A_2}{A_1}$$

where $$\frac{A_2}{A_1} = \frac{[(1-\Phi)+ZM+(a'+b')M]}{[Z+(a+b)(1-\Phi)]}$$

PROB. 9-16 Const. Optimum

The objective function in this expression is xy and the constaint is $x^2 + y^2 = 10$ (this relates the values of x & y).

$$\therefore x^2 + y^2 = 10 \Rightarrow y = (10 - x^2)^{0.5}$$

Therefore, the objective function

$$\Phi(x,y) = xy = x(10 - x^2)^{0.5} = \phi(x)$$

$$\Rightarrow \left.\frac{d\phi}{dx}\right|_{x_{opt}} = 0 = \frac{x_{opt}}{\cancel{2}}(10 - x_{opt}^2)^{-0.5}(-\cancel{2}x_{opt}) + (10 - x_{opt}^2)^{0.5}$$

Solving for x_{opt}, we have $x_{opt} = 0\sqrt{5})^{0.5}$

& $y_{opt} = \sqrt{5}$

where only $x \geqq 0$ and $y \geqq$ are being sought.

PROB. 9-17 Const Optimum Using Lagrange Multiplier

Objective Function $u(x,y,z) = x + 2y^2 + z^2$

Constraint $v(x,y,z) = x + y + z - 1$

Forming the Lagrange expression

$$L(x,y,z,\lambda) = x + 2y^2 + z^2 + \lambda(x + y + z - 1)$$

To find the optimum, we set the partial derivative of L with respect to $x, y, z,$ & $\lambda = 0$ and solve

$$\therefore \frac{\partial L}{\partial x} = 1 + \lambda = 0 \implies \lambda = -1$$

$$\frac{\partial L}{\partial y} = 4y + \lambda = 0 \implies y = -\frac{1}{4}\lambda = 0.25$$

PROB. 9-17 (CON'T)

$$\frac{\partial L}{\partial z} = 2z+\lambda=0 \Rightarrow z = -\frac{1}{2}\lambda = 0.5$$

$$\frac{\partial L}{\partial \lambda} = x+y+z-1=0 \Rightarrow x = 1-y-z = 0.25$$

$$\therefore \boxed{x_{opt} = 0.25 \, ; \, y_{opt} = 0.25 \, ; \, z_{opt} = 0.5}$$

PROB. 9-18

CASE A: No inflation

ALT. 1: Present Value = $5800

ALT. 2: Present Value = $5000 + 2500(1+0.08)^{-8} = \$6351$

∴ SELECT ALTERNATIVE 1 AS IT HAS LOWEST PRESENT VALUE

CASE B: 7% INFLATION ($2500 IS COST 8 YEARS FROM NOW)

ALT. 1: Present Value = $5800

ALT. 2: $PV = 5000 + \left[(2500)(1+0.07)^{-8}\right](1+0.08)^{-8}$

$= \$5786$

∴ SELECT ALTERNATIVE 2 AS IT HAS LOWEST PRESENT VALUE

(NOTE: INFLATION & COST OF CAPITAL SHOULD BE CONSIDERED SEPARATELY, I.E.,

$(1+0.07)^{-8}(1+0.08)^{-8} \neq (1+0.15)^{-8}$!)

Problem 9-19 Optimum ΔT_{min}(Using Example 9-8) Network

Input Data: $\Phi := 0.35$ $MAR := 0.15$

Stream Data:

Stream S1

$CP\_S1 := 50\ \frac{kJ}{s \cdot C}$

$In\_S1\_spec := 20\ C$

$Out\_S1\_spec := 160\ C$

$x_{S1a} := 0.48$

$CP\_S1a := x_{S1a} \cdot CP\_S1$

$CP\_S1b := \left(1 - x_{S1a}\right) \cdot CP\_S1$

Stream S2

$CP\_S2 := 55\ \frac{kJ}{s \cdot C}$

$In\_S2\_spec := 120\ C$

$Out\_S2\_spec := 260\ C$

Stream S3

$CP\_S3 := 30\ \frac{kJ}{s \cdot C}$

$In\_S3\_spec := 280\ C$

$Out\_S3\_spec := 60\ C$

Stream S4

$CP\_S4 := 40\ \frac{kJ}{s \cdot C}$

$In\_S4\_spec := 180\ C$

$Out\_S4\_spec := 20\ C$

Hot Utility

$T\_HOS := 320\ C$

$T\_HOR := 310\ C$

$C_{HO} := 2.25 \cdot 10^{-6}$ \$/kJ

Cold Utility

$T\_CWS := 10\ C$

$T\_CWR := 20\ C$

$C_{CW} := 0.25 \cdot 10^{-6}$ \$/kJ

Exchanger Capital Cost Estimate

$Cost\_Exchr(A) := 1000 \cdot A^{0.6}$ where A is the surface area in m²

Exchanger Overall Heat Transfer Coefficients

$U_{E1} := 1 \frac{W}{m^2 \cdot s}$ $U_{E2} := 1 \frac{W}{m^2 \cdot s}$ $U_{E3} := 1 \frac{W}{m^2 \cdot s}$ $U_{E4} := 1 \frac{W}{m^2 \cdot s}$

$U_{E5} := 1 \frac{W}{m^2 \cdot s}$ $U_{C1} := 1 \frac{W}{m^2 \cdot s}$ $U_{H1} := 1 \frac{W}{m^2 \cdot s}$

Set up calculations for sequence of ΔT_{min}

Pinch Information

$T_{pinch} := 120\ C$ $klast := 81$ $DT_{first} := 2$ $DT_{last} := 22$

$k := 1, 2 .. klast$

$\delta := \frac{\left(DT_{last} - DT_{first}\right)}{klast - 1}$ $\Delta T_{min_k} := DT_{first} + (k - 1) \cdot \delta$ *Sequence of ΔT_{min}*

Define Functions Useful In Calculations

$\mathrm{QDOT}(CP, \Delta T) := CP \cdot \Delta T$ Heat Rate

$\Delta T_{LM}(\Delta T1, \Delta T2) := \dfrac{(\Delta T1 - \Delta T2)}{\ln\left(\dfrac{\Delta T1}{\Delta T2}\right)}$ log-mean temperature difference

$\mathrm{Area}(\mathrm{QDOT}, U, \mathrm{LMTD}) := \dfrac{\mathrm{QDOT}}{U \cdot \mathrm{LMTD}}$ Exchanger Surface Area

For network on p.432 - starting at the pinch - calculations repeated for each k

Exchanger E-3

$\mathrm{In\_S4E3} := \mathrm{In\_S4\_spec}$

$\mathrm{Out\_S4E3}_k := T_{pinch} + \Delta T_{min_k}$

$\mathrm{In\_S2E3} := \mathrm{In\_S2\_spec}$

$\Delta T_k := \mathrm{In\_S4E3} - \mathrm{Out\_S4E3}_k$

$DQ_{E3_k} := \mathrm{QDOT}(\mathrm{CP\_S4}, \Delta T_k)$ $\frac{kJ}{s}$ Heat Rate

$\mathrm{Out\_S2E3}_k := \mathrm{In\_S2E3} + \dfrac{DQ_{E3_k}}{\mathrm{CP\_S2}}$

$\Delta T1_k := \mathrm{In\_S4E3} - \mathrm{Out\_S2E3}_k$

$\Delta T2_k := \mathrm{Out\_S4E3}_k - \mathrm{In\_S2E3}$

$\mathrm{LMTD\_E3}_k := \Delta T_{LM}(\Delta T1_k, \Delta T2_k)$

$A_{E3_k} := \mathrm{Area}\left(DQ_{E3_k}, U_{E3}, \mathrm{LMTD\_E3}_k\right)$

$C_{E3_k} := \mathrm{Cost\_Exchr}\left(A_{E3_k}\right)$ \$

Exchanger E-2

$\mathrm{In\_S4E2}_k := \mathrm{Out\_S4E3}_k$

$\mathrm{In\_S1bE2} := \mathrm{In\_S1\_spec}$

$\mathrm{Out\_S1bE2} := T_{pinch}$

$\Delta T := \mathrm{Out\_S1bE2} - \mathrm{In\_S1bE2}$

$DQ_{E2} := \mathrm{QDOT}(\mathrm{CP\_S1b}, \Delta T)$ $\frac{kJ}{s}$ Heat Rate

$\mathrm{Out\_S4E2}_k := \mathrm{In\_S4E2}_k - \dfrac{DQ_{E2}}{\mathrm{CP\_S4}}$

$\Delta T1_k := \mathrm{In\_S4E2}_k - \mathrm{Out\_S1bE2}$

$\Delta T2_k := Out\_S4E2_k - In\_S1bE2$

$LMTD\_E2_k := \Delta T_{LM}(\Delta T1_k, \Delta T2_k)$

$A_{E2_k} := Area(DQ_{E2}, U_{E2}, LMTD\_E2_k)$ m^2

$C_{E2_k} := Cost\_Exchr(A_{E2_k})$ $

Exchanger E-4

$Out\_S1E4 := Out\_S1\_spec$

$In\_S1E4 := T_{pinch}$

$Out\_S3E4_k := T_{pinch} + \Delta T_{min_k}$

$\Delta T := Out\_S1E4 - In\_S1E4$

$DQ_{E4} := QDOT(CP\_S1, \Delta T)$ $\frac{kJ}{s}$ Heat Rate

$In\_S3E4_k := Out\_S3E4_k + \frac{DQ_{E4}}{CP\_S3}$

$\Delta T1_k := In\_S3E4_k - Out\_S1E4$

$\Delta T2_k := Out\_S3E4_k - In\_S1E4$

$LMTD\_E4_k := \Delta T_{LM}(\Delta T1_k, \Delta T2_k)$

$A_{E4_k} := Area(DQ_{E4}, U_{E4}, LMTD\_E4_k)$ m^2

$C_{E4_k} := Cost\_Exchr(A_{E4_k})$ $

Exchanger E-1

$In\_S3E1_k := Out\_S3E4_k$

$Out\_S3E1 := Out\_S3\_spec$

$In\_S1aE1 := In\_S1\_spec$

$Out\_S1aE1 := Out\_S1bE2$

$\Delta T_k := In\_S3E1_k - Out\_S3E1$

$DQ_{E1_k} := QDOT(CP\_S3, \Delta T_k)$ $\frac{kJ}{s}$ Heat Rate

$\Delta T1_k := In\_S3E1_k - Out\_S1aE1$

$\Delta T2 := Out\_S3E1 - In\_S1aE1$

$\mathrm{LMTD\_E1}_k := \Delta T_{LM}(\Delta T1_k, \Delta T2)$

$A_{E1_k} := \mathrm{Area}(DQ_{E1_k}, U_{E1}, \mathrm{LMTD\_E1}_k)$ m^2

$C_{E1_k} := \mathrm{Cost\_Exchr}(A_{E1_k})$ \$

Exchanger E-5

$\mathrm{In\_S2E5}_k := \mathrm{Out\_S2E3}_k$

$\mathrm{In\_S3E5} := \mathrm{In\_S3\_spec}$

$\mathrm{Out\_S3E5}_k := \mathrm{In\_S3E4}_k$

$\Delta T_k := \mathrm{In\_S3E5} - \mathrm{Out\_S3E5}_k$

$DQ_{E5_k} := \mathrm{QDOT}(\mathrm{CP\_S3}, \Delta T_k)$ $\frac{kJ}{s}$ Heat Rate

$\mathrm{Out\_S2E5}_k := \mathrm{In\_S2E5}_k + \frac{DQ_{E5_k}}{\mathrm{CP\_S2}}$

$\Delta T1_k := \mathrm{In\_S3E5} - \mathrm{Out\_S2E5}_k$

$\Delta T2_k := \mathrm{Out\_S3E5}_k - \mathrm{In\_S2E5}_k$

$\mathrm{LMTD\_E5}_k := \Delta T_{LM}(\Delta T1_k, \Delta T2_k)$

$A_{E5_k} := \mathrm{Area}(DQ_{E5_k}, U_{E5}, \mathrm{LMTD\_E5}_k)$ m^2

$C_{E5_k} := \mathrm{Cost\_Exchr}(A_{E5_k})$ \$

Exchanger C-1

$\mathrm{In\_CWC1} := \mathrm{T\_CWS}$

$\mathrm{Out\_CWC1} := \mathrm{T\_CWR}$

$\mathrm{In\_S4C1}_k := \mathrm{Out\_S4E2}_k$

$\mathrm{Out\_S4C1} := \mathrm{Out\_S4\_spec}$

$\Delta T_k := \mathrm{In\_S4C1}_k - \mathrm{Out\_S4C1}$

$DQ_{C1_k} := \mathrm{QDOT}(\mathrm{CP\_S4}, \Delta T_k)$ $\frac{kJ}{s}$ Heat Rate

$\Delta T1_k := \mathrm{In\_S4C1}_k - \mathrm{Out\_CWC1}$

$\Delta T2 := \mathrm{Out\_S4C1} - \mathrm{In\_CWC1}$

$\mathrm{LMTD\_C1}_k := \Delta T_{LM}(\Delta T1_k, \Delta T2)$

$A_{C1_k} := \mathrm{Area}\left(DQ_{C1_k}, U_{C1}, \mathrm{LMTD\_C1}_k\right)$ m^2

$C_{C1_k} := \mathrm{Cost\_Exchr}\left(A_{C1_k}\right)$ $

Exchanger H-1

$\mathrm{In\_S2H1}_k := \mathrm{Out\_S2E5}_k$

$\mathrm{Out\_S2H1} := \mathrm{Out\_S2\_spec}$

$\mathrm{In\_HOH1} := \mathrm{T\_HOS}$

$\mathrm{Out\_HOH1} := \mathrm{T\_HOR}$

$\Delta T_k := \mathrm{Out\_S2H1} - \mathrm{In\_S2H1}_k$

$DQ_{H1_k} := \mathrm{QDOT}\left(\mathrm{CP\_S2}, \Delta T_k\right)$ $\frac{kJ}{s}$ Heat Rate

$\Delta T1_k := \mathrm{Out\_HOH1} - \mathrm{In\_S2H1}_k$

$\Delta T2 := \mathrm{In\_HOH1} - \mathrm{Out\_S2H1}$

$\mathrm{LMTD\_H1}_k := \Delta T_{LM}\left(\Delta T1_k, \Delta T2\right)$

$A_{H1_k} := \mathrm{Area}\left(DQ_{H1_k}, U_{H1}, \mathrm{LMTD\_H1}_k\right)$ m^2

$C_{H1_k} := \mathrm{Cost\_Exchr}\left(A_{H1_k}\right)$ $

Cost Summary

$\mathrm{Cost\_Hot}_k := DQ_{H1_k} \cdot C_{HO} \cdot 3600 \cdot 24 \cdot 365$ $/y

$\mathrm{Cost\_Cold}_k := DQ_{C1_k} \cdot C_{CW} \cdot 3600 \cdot 24 \cdot 365$ $/y

$\mathrm{Cost\_Util}_k := \mathrm{Cost\_Hot}_k + \mathrm{Cost\_Cold}_k$

$\mathrm{Cap\_Exchr}_k := C_{E1_k} + C_{E2_k} + C_{E3_k} + C_{E4_k} + C_{E5_k} + C_{C1_k} + C_{H1_k}$ $

$T_{hs_k} := 5 \cdot \mathrm{Cap\_Exchr}_k$ $ Estimated Total Investment Cost

$\mathrm{Ann\_Depr}_k := \frac{\mathrm{Cap\_Exchr}_k}{7}$ $/y

$C_{T_k} := \mathrm{Cost\_Util}_k \cdot (1 - \Phi) + \mathrm{Ann\_Depr}_k \cdot (1 - \Phi) + \mathrm{MAR} \cdot T_{hs_k}$ $/y**Total Annual Cost**

Construct Table of Answers and Plot

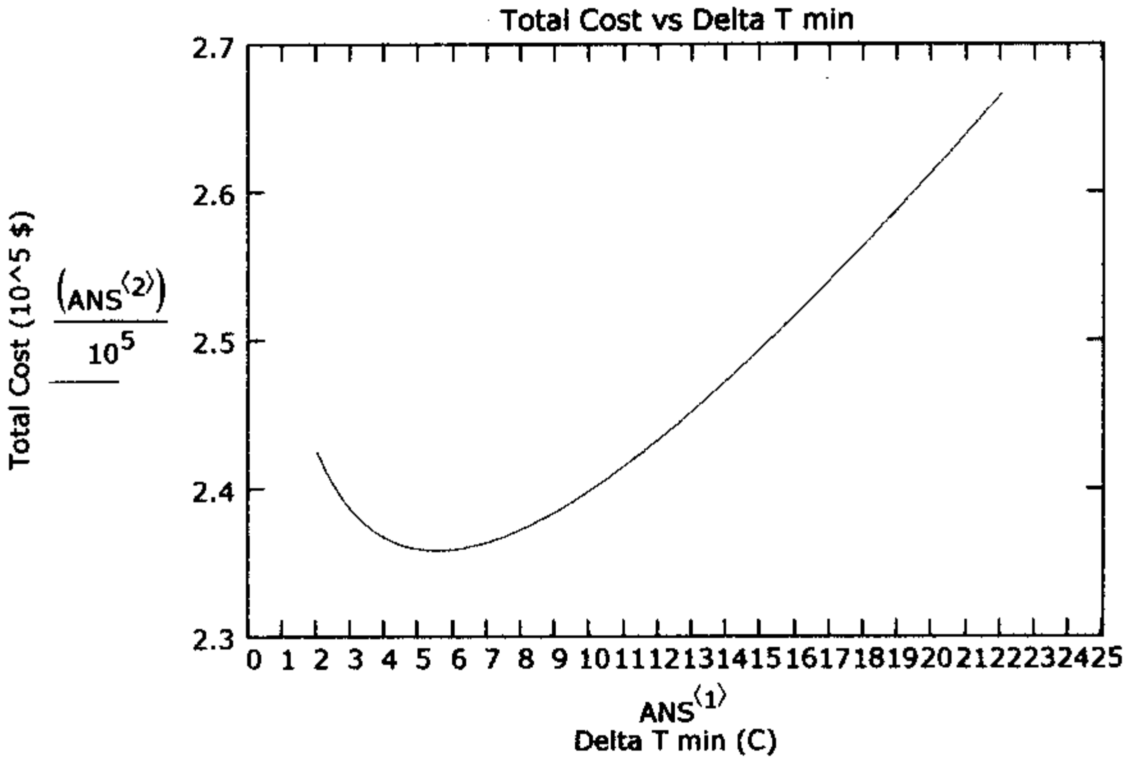
$ANS_{k,1} := \Delta T_{min_k}$

$ANS_{k,2} := C_{T_k}$

$ANS_{k,3} := Cost\_Util_k$

ANS =

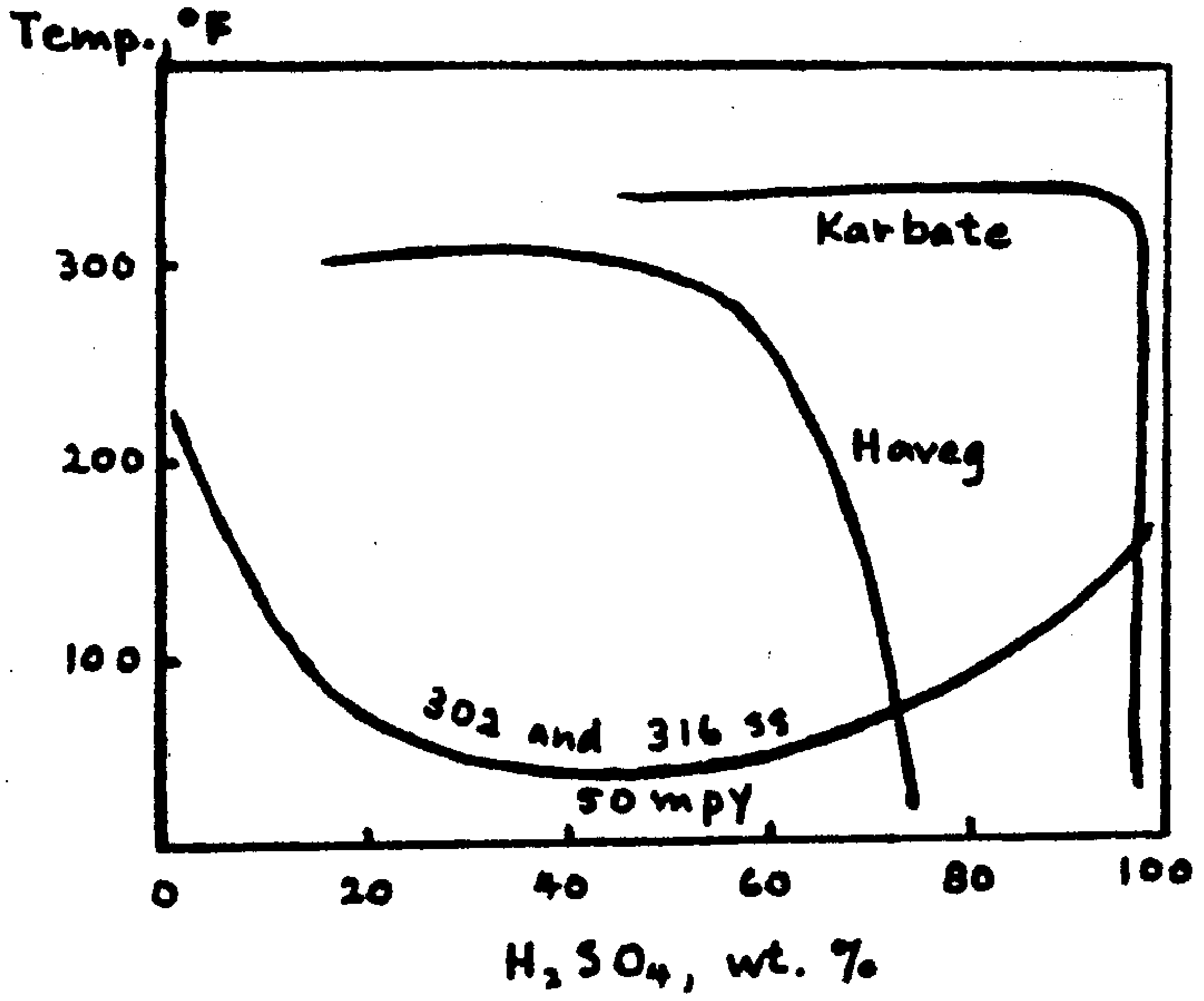
	ΔT_{min}	C_T	C_{UTIL}
	1	2	3
1	2	$2.425 \cdot 10^5$	$1.99 \cdot 10^5$
2	2.25	$2.412 \cdot 10^5$	$2.003 \cdot 10^5$
3	2.5	$2.402 \cdot 10^5$	$2.016 \cdot 10^5$
4	2.75	$2.393 \cdot 10^5$	$2.03 \cdot 10^5$
5	3	$2.386 \cdot 10^5$	$2.043 \cdot 10^5$
6	3.25	$2.379 \cdot 10^5$	$2.056 \cdot 10^5$
7	3.5	$2.374 \cdot 10^5$	$2.069 \cdot 10^5$
8	3.75	$2.37 \cdot 10^5$	$2.082 \cdot 10^5$
9	4	$2.366 \cdot 10^5$	$2.096 \cdot 10^5$
10	4.25	$2.364 \cdot 10^5$	$2.109 \cdot 10^5$
11	4.5	$2.362 \cdot 10^5$	$2.122 \cdot 10^5$
12	4.75	$2.36 \cdot 10^5$	$2.135 \cdot 10^5$
13	5	$2.359 \cdot 10^5$	$2.148 \cdot 10^5$
14	5.25	$2.358 \cdot 10^5$	$2.162 \cdot 10^5$
15	5.5	$2.358 \cdot 10^5$	$2.175 \cdot 10^5$
16	5.75	$2.358 \cdot 10^5$	$2.188 \cdot 10^5$

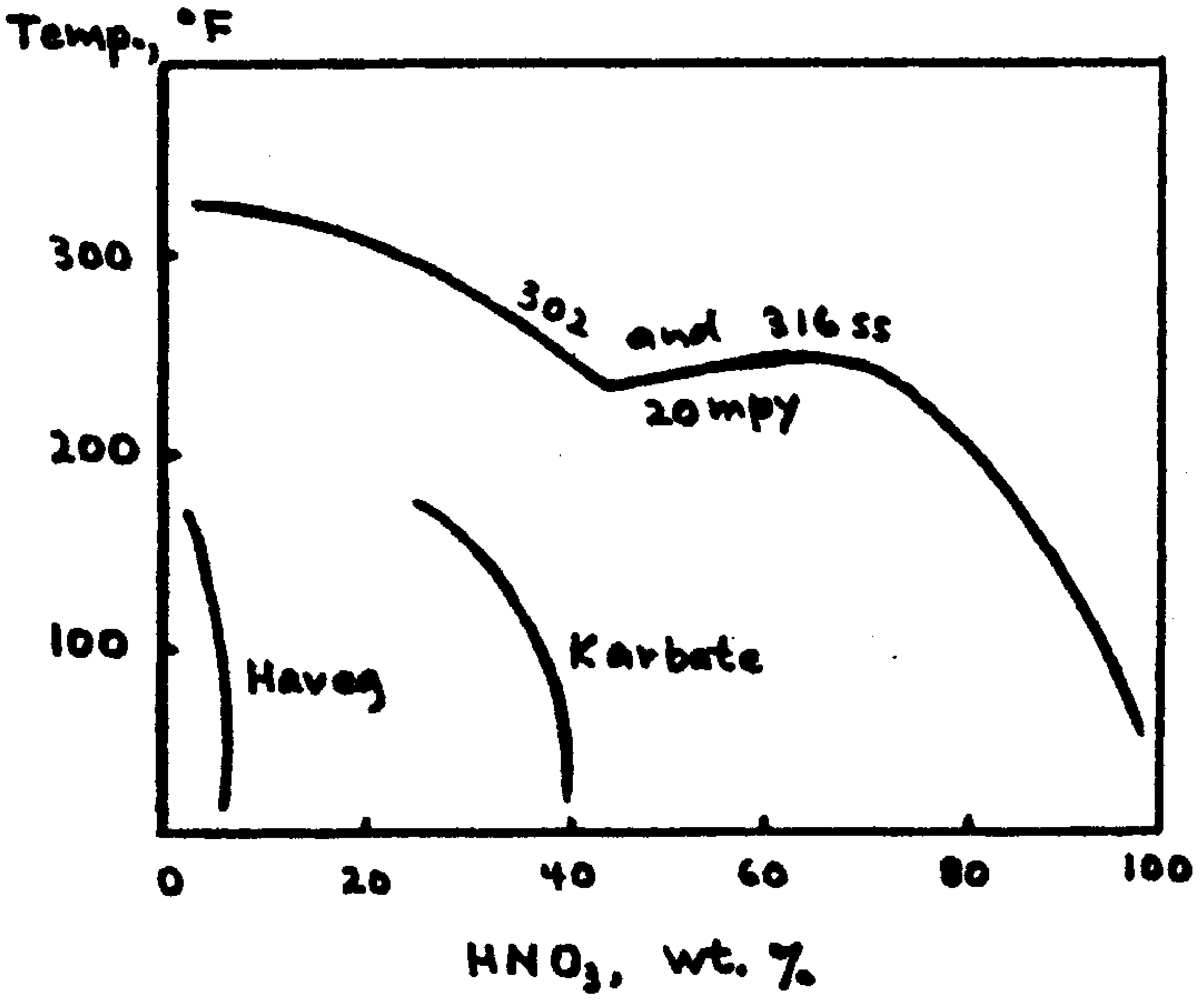


Very flat optimum between ΔT_{min} of 5 to 6 degrees C. Largest exchanger area is 181 m^2.

10-1

Approximate plots showing conditions of generally acceptable corrosion resistance can be constructed from information available in "Perry's Chemical Engineers' Handbook" and regular materials review including corrosion resistance published by Chemical Engineering.





mpy = mils per year

10-2

Allowable corrosion is 3.175 mm or 0.003175 m. Assume interest payable at end of year

	Investment comparison		
	Ni-clad	Monel-clad	Hastelloy
Total installed cost, $	88,000	105,000	198,000
Additional cost over Ni-clad, $		17,000	110,000
Estimated life, y	6.35	12.7	27.9
Replacement cost, $	79,200	94,500	178,200
Annual replacement cost, $ (replacement cost/est. life)	12,470	7,440	6,390
Annual interest charge, $	8,800	10,500	19,800
Total annual cost, $	21,270	17,940	26,190
Annual savings over Ni-clad	-	3,330	NA
Return on investment, % (annual savings/add. cos) 100	-	19.6	NA

The most suitable material of construction for the kettle is Monel-clad.

Answer

10-3

The additional information necessary in the design of the rotary vacuum filter includes values of α, β, μ, and w. (See Chapter 15 of text for nomenclature.) Added information needed to design rotary filters includes: T, Δp, ψ_a, ψ_f, safety factor, vacuum pump efficiency, and N_r.

From laboratory tests with the above data, we can calculate α and β (μ and w can be obtained from lab tests). Thus all the available information is available to calculate the area of the rotary filter under the final conditions using Eq. (15-101). Equations (12-22) to (12-25) and Eqs. (15-102) to (15-110) can be used to calculate the size of the vacuum pump that is required.

Answer

10-4

The density of 316ss is 7850 kg/m^3

The 316ss will only be available as a square sheet that will need to be cut into a circle with a diameter of 0.5 m.

$$m = (0.5)^2(0.05)(7850) = 98.1 \text{ kg}$$

Assume machined material is 50% of purchased cost

$$\text{total cost} = (98.1)(2)(\$10.00) = \underline{\$1960}$$

Answer

Alternate method involves estimation of fabrication time

cutting and trimming	6 h
drilling holes	8 h
finishing	8 h
Total hours	22 h

Labor at \$25/h plus 50% overhead added to material cost

$$\text{total cost} = (98.1)(\$10.00) + (22)(1.5)(\$25.00)$$
$$= \underline{\underline{\$1810}}$$

Alternate Answer

10-5

The selection of materials for the available tanks is shown below.

Tanks	Material
Brass-lined	Water
Carbon steel	98% sulfuric acid
Concrete	20% hydrochloric acid
Nickel-lined	10% caustic soda
316 stainless	75% phosphoric acid
Wood	Vinegar

Answer

10-6

Information on cylindrical tank is the following:

$D_i = 3.6 m$ $L = 11 m$ $E = 0.9$ $P = 790 kPa = 690 kPag$

$T = 25°C$

From Table 12-10 for a cylindrical shell

$$t = \frac{P r_i}{S E_J - 0.6P} + C_c \qquad \text{when } t \le r_i/2$$

S is the maximum allowable working stress in kPa. It is often assumed that this value is essentially one fourth of the ultimate tensile strength at the operating temperature. "Perry's Chemical Engineer's Handbook.", 6th ed., p 23-41 provides an ultimate tensile strength of 62,400 psia or 430,000 kPa:

Assume that C_c is negligible relative to the first term

$$t = \frac{(690)(1.8)}{(430{,}000/4)(0.9) - 0.6(690)}$$

$= 0.0128$ or $\underline{\underline{0.013 m}}$ to take care of C_c

Answer

10-7

This is a standard type of problem treated in any reference dealing with nuclear power reactor design. There are also numerous examples of this type of treatment in the literature including J. R. Envico, "Corrosion of Nuclear Materials," Materials 7(10), 464 (1967) and R. E. Skavdahl and E. L. Zebroski, "Finding Materials for Fast Reactors in the Future," Chem. Eng., 76 (17), 114 (1969). Other reference materials are available in handbook published on this subject.

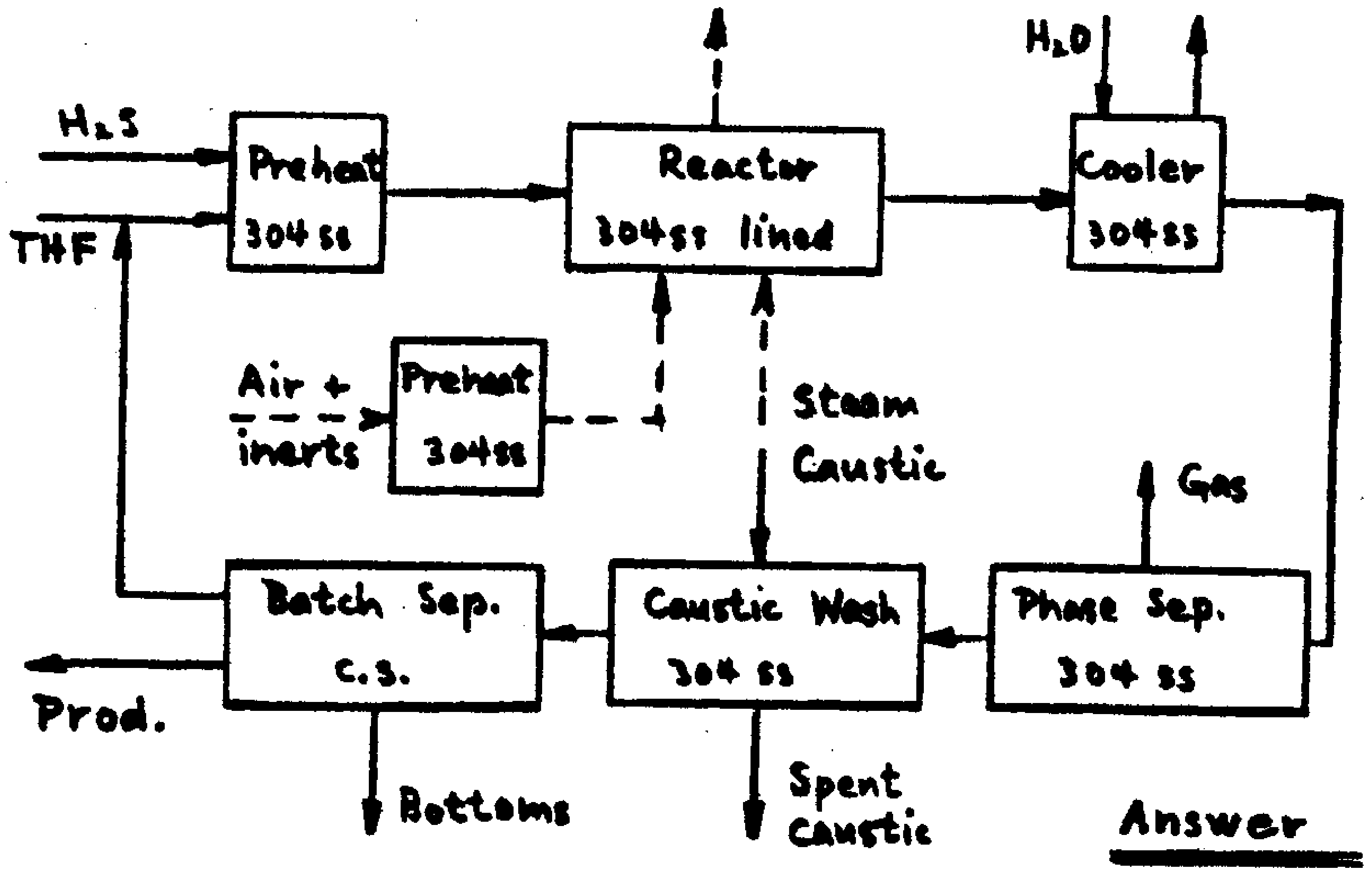
Typical data or information needed for determining whether a given tube diameter and gauge number is satisfactory are given below:

- Type of stainless steel
    - Poisson's ratio
    - Young's modulus
    - Coefficient of thermal expansion
    - Thermal conductivity
    - Creep stress limit
- Type of heating
- Rate of heat transfer
- Properties of fluids for heat transfer calculations
- Temperature at wall surfaces and wall ΔT
- Number of tubes
- Length of tubes
- Safety factors to be considered

Answer

10-8

The general flowsheet for thiophane production is given below. Materials of construction for each unit are included.



Answer

10-9

Two systems were developed in Problem 2-13 to transfer liquid chlorine from a liquid storage container; one using recompressed chlorine vapor and the other using vaporized chlorine. Flow diagrams describe these systems. Please refer back to Problem 2-13 for details. The information below applies to either system.

For dry chlorine gas, 317ss, nickel, Monel, and carbon steel have been the recommended materials of construction. In the event of water contamination of the chlorine gas, carbon steel should not be considered. If there is a potential for leakage of the chlorine, the most acceptable material of construction would be 317ss.

If wet chlorine is involved, other choices, besides 317ss, are Nucerite (a ceramic-metal composite), glass-clad steel or Haveg. The least expensive of these choices generally is 317ss because of its greater flexibity to meet special design conditions.

<u>Answer</u>

10-10

With reference to Fig. 2-1 in the 4th ed. of Plant Design and Economics for Chemical Engineers, each piece of equipment identified in the nitric acid plant is listed below with the most appropriate material of construction. Note that other choices are possible but have not been included because they are generally more expensive.

Piece of equiment	Material choice	
Ammonia vaporizer	cs	
Ammonia filter	cs	
Air filter	cs	
Oxidation chamber	430 ss	850°C top temp.
Cooler condenser (lower)	430 ss	
Cooler condenser (upper)	430 ss	
Absorption tower	430 ss	
Converter	316 ss	
Mixing chamber	316 ss	
Preheater	316 ss	
Reheater	316 ss	

<u>Answer</u>

11-1

Depending upon the type of report prepared, company guidelines for preparing written reports can vary significantly from the typical outline presented in the chapter. If the report is prepared for use within the design group, it may eliminate the letter of transmittal, title page, and table of contents (particularly if the report is not lengthy.

The body of the report also can be decreased when reporting to the design group who are familiar with the project. For example, the report could provide the recommended design and provide sufficient economic details that strongly support or reject the project. If the report is prepared for the decision makers in the company, greater emphasis will probably be required if investment in the project is recommended. Such a report will generally have a much shorter appendix since the decision makers will not be interested in the calculation details, but how investment in the project will affect the bottom line of the company. The decision makers will assume that the design group has carefully reviewed the design calculations and have made a sensitivity analysis of the consequences of their recommendation before its transmittal.

11-2

Chemical engineering department guidelines for preparing laboratory reports generally follow the typical outline for process design projects as presented in this chapter. Sometimes the departmental guidelines request more information in the body of the report to provide immediate access to the assumptions made and the justification for those assumptions. Also, often more information is required in the analysis of possible sources of error and their potential effect on the final result. Some reports may not require an economic evaluation particularly when only experimental results are being analyzed. In fact, there may be several different outlines depending on the purpose of the report.

11-3

Review of a design report prepared by a colleague will need to be done by the reviewer. Answering the questions posed in in the problem should provide the reviewer with positive suggestions that can be transmitted back to the colleague.

11-4

Every design report must be checked with a spell checker to avoid embarrestment on the part of the writer, but also to emphasize to the reader that the writer is a professional who has made every effort to prepare a technical study of the highest quality.

11-5

There are numerous software packages that prepare various types of visual material. These packages can cover the preparation of simple graphs to the delopment of highly automated visuals. The cost for the packages that prepare simple but instructive graphs are quite modest and can be programmed to provide various ways to display the material of interest using color to provide desired hightlights. The more complex packages are considerably more expensive, but the cost has decreased dramatically in the past decade. Such software packages are capable of displaying animated sequences, halftones, and videos. Because of the competition, these software packages are continuously being improved with greater flexibility for the user. With the constant updating of these software packages it is difficult to specify advantages and disadvantages of a specific package since these change even over a short time period.
Regardless of the software selected, there are numerous guidelines concerning the density of information that is to be used in visuals. The font size of the lettering needs to be adjusted to meet the requirements placed on it by the location of the presentation. Colors on the visual need to be selected that will enhance the images projected. Whatever material that is to be presented should be easily readable by all those attending the presentation.

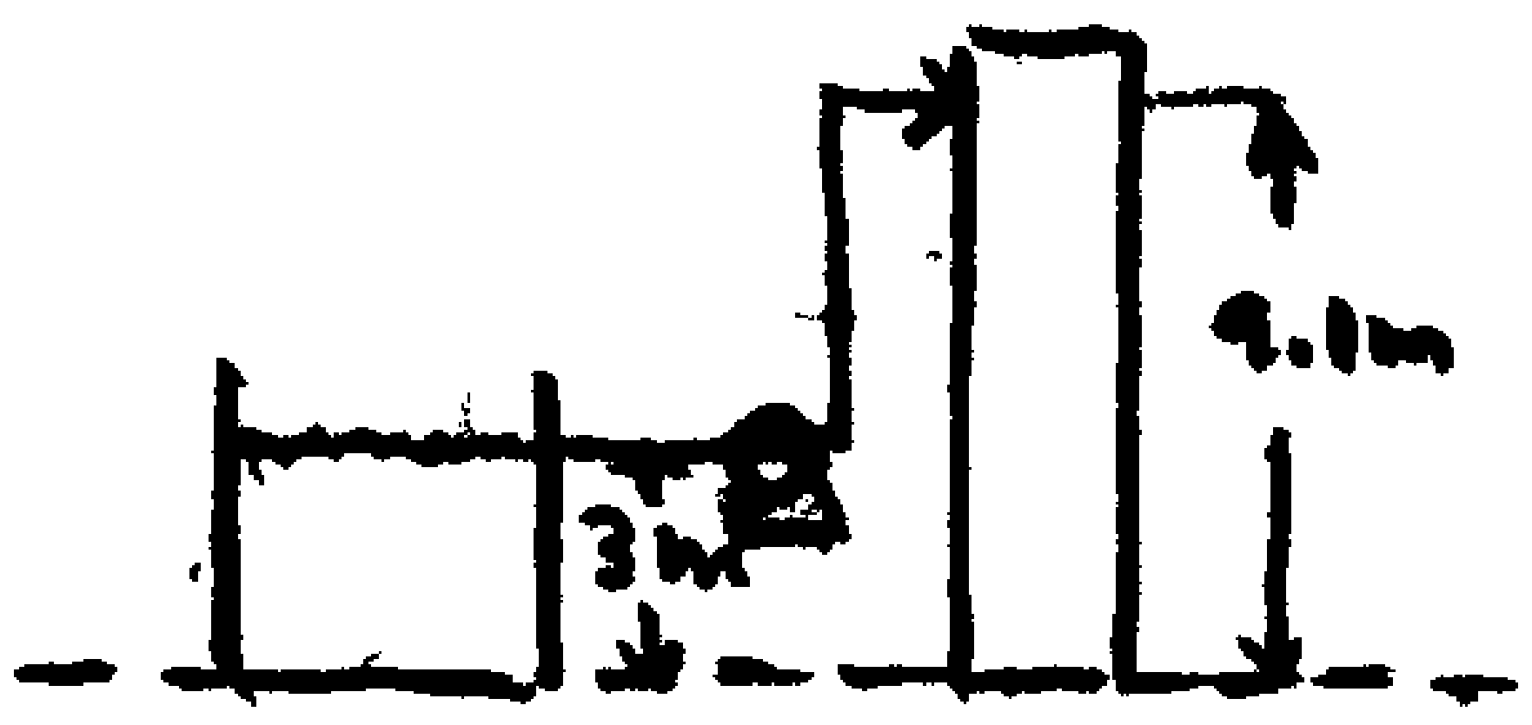
11-6

A skeleton outline and a detailed outline for a final report on the detailed-estimate design of a distillation unit can have many forms. The skeleton outline will probably include the items listed.

1. Introduction to the project
   Listing of items specifically requested
   Actions taken to meet those requests
2. Results from the project
3. Recommended design
   Brief description of proposed design
   Listing of economics supporting recommendation
4. Conclusions

The detailed outline will be very similar to the outline detailed in Table 11-1. However, in the Discussion section (V.C.) that covers the technical matters of importance there will be a detailed analysis of the steps involved in the deisign of a distillation as outlined in Fig. 15-1 of the text. This analysis will recognize the specific problems that are encountered in obtaining high purity methanol from a water-methanol mixture.

12-1



$D_i = 0.078\,m$ $\quad A_i = \pi D_i^2/4 = 4.778 \times 10^{-3}\,m^2$

5-90° elbows

$\dot{m}_i = 2.7\,kg/s$

$\mu_i = 15\,cP$

$\rho_i = 857\,kg/m^3$

$\eta_p = 0.4$

Find power requirement for pump

$$g z_1 + p_1 v_1 + V_1^2/2\alpha + \int p\,dv + W = g z_2 + p_2 v_2 + V_2^2/2\alpha + F$$

$$V = 2.7\,\frac{kg}{s}\;\frac{m^3}{857\,kg}\;\frac{1}{4.778 \times 10^{-3}\,m^2} = 0.659\,m/s$$

$$Re = DV\rho/\mu = (0.078)(0.659)(857)/0.015 = 2936$$

From Fig. 12-1, $\varepsilon/D = 0.000046/0.078 = 0.00059$ based on commercial steel, $f = 0.011$

$$L_e = 46 + 5(32)(0.078) = 58.5\,m$$

$$F = 2 f V^2 L_e/D$$
$$= (2)(0.011)(0.659)^2(58.5)/0.078 = 7.17$$

Assume $v_1 = v_2$, $\int p\,dv = 0$ and $V_1 = 0$

$$W = g(z_2 - z_1) + (p_2 - p_1)/\rho + V_2/2 + F$$
$$= (9.806)(6.1) + (445 - 101)/857 + (0.659)^2/2 + 7.17$$
$$\doteq 468.6\,Nm/kg$$

$$P_{theo} = (468.6)(2.7) = 1265\,W = 1.27\,kW$$

$$P_{act} = (1.265)/0.4 = \underline{\underline{3.18\,kW}} \text{ or } 3.2\,kW$$

Answer

12-2

$\dot{m} = 2.14$ kgs

$L = 21$ m

$D_i = 0.0409$ m

$\mu_L = 5 \times 10^{-4}$ Pa.s

$\rho_L = 849$ kg/m^3

The pressure drop is entirely from friction since there is no elevation change. Equivalent resistances of valves and fittings are obtained from Table 12-1.

	Fittings and valves	Equivalent resistance in pipe diameters
6	90° elbows	192
1	tee (used as elbow)	60
1	gate valve (open)	7
1	globe valve (open)	300
		559

Equivalent length = (559)(0.0409) = 22.9 m

Total length of pipe = 21 + 22.9 = 43.9 m

$V_i = (2.14)/(849)(0.785)(0.0409)^2 = 1.92$ m/s

$Re = DV\rho/\mu = (0.0409)(1.92)(849)/0.0005 = 1.33 \times 10^5$

$\epsilon/D_i = 0.000046/0.0409 = 0.0011$

$f = 0.0053$ from Fig. 12-1

$F = 2fV^2L_e/D = (2)(0.0053)(1.92)^2(43.9)/0.0408$

$= 41.9$ Nm/kg

$\Delta p = F\rho = (41.9)(849) = 35{,}575$ Pa $= \underline{\underline{35.6 \text{ kPa}}}$

Answer

12-3

Volumetric flow rate of cooling water = 0.0095 m^3/s

$\rho_L = 996\ kg/m^3$ $\quad \mu_L = 0.893\ cP$

$\dot{m}_{H_2O} = (0.0095)(996) = 9.46\ kg/s$

Assume that Re > 2100, use Eq.(12-15) to determine optimum pipe diameter

$$D_{i,opt} = 0.344\ \dot{m}_v^{0.45}\ \rho^{0.13} \quad \text{where } D_i \text{ is in meters}$$

$$= (0.344)(0.0095)^{0.45}\ (996)^{0.13} = 0.1038\ m = 4.087\ in.$$

Check Re,

$$Re = DG/\mu = (0.1038)(9.46/0.785(0.1038)^2/0.000893$$

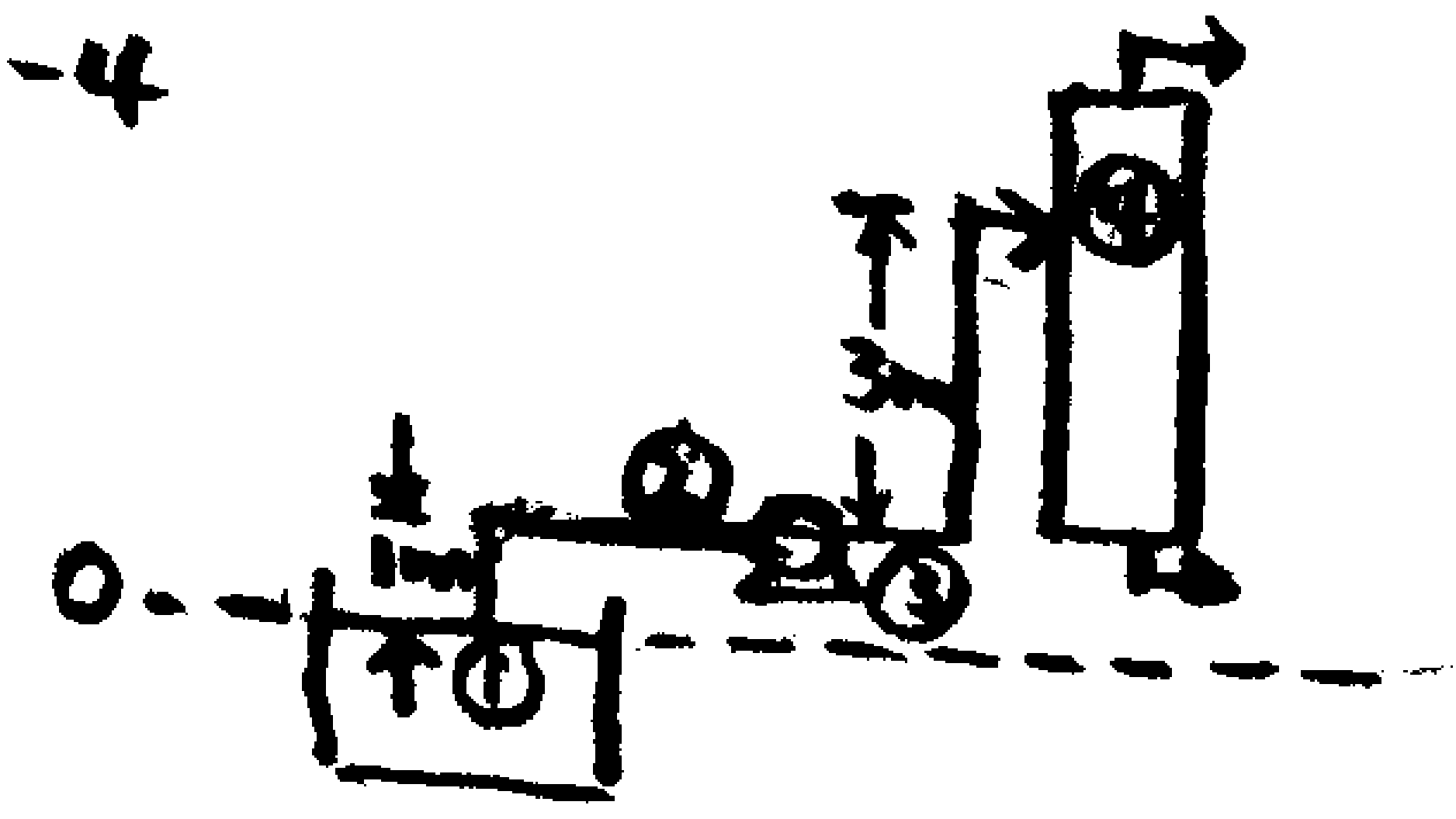
$$= 130{,}000 \quad \therefore Re > 2100$$

Thus, the use of Eq.(12-15) is valid.

The inside diameter of a Schedule 40, 4 in. steel pipe is 4.026 in. The next larger size is 5.047 in.. Therefore, a 4 in., Schedule 40 pipe should be selected even though it is slightly smaller than the computed value of 4.087 in.

Answer

12-4



$T_{82} = 38°C$

$\overset{\circ}{p}_{v82} = 26.4 \, kN/m^2$

$\rho_{82} = 860 \, kg/m^3$

$\dot{m}_v = 0.0025 \, m^3/s$

$\eta_p = 0.60$

$p_1 = 101.3 \, kPa$

$\Delta p_{1-2} = 3450 \, N/m^2$

$\Delta p_{3-4} = 37{,}900 \, N/m^2$

Assume Re > 2100, use Eq. (12-15) to obtain optimum pipe diam.

$$D_{i,opt} = 0.344 \, \dot{m}_v^{0.45} \rho^{0.13} = (0.344)(0.0025)^{0.45}(860)^{0.13} = 0.0559 \, m$$

Use a 2in., Schedule 40 (2.067 in or 0.0525 m) pipe for the system.

$$V_4 = (0.0025)/(0.785)(0.0525)^2 = 1.15 \, m/s$$

Designate the level of the tank as the datum plane for the pumping system. Since $V_1 = 0$, the mechanical energy balance, Eq. (12-2) can be rearranged into

$$W\eta_p = p_4/\rho + g Z_4 + V_4^2/2\alpha - p_1/\rho + F \quad \text{where } \alpha = 0$$

$$= \frac{445{,}000}{860} + (9.806)(4) + \frac{(1.15)^2}{2} - \frac{101{,}300}{860} + \frac{3450 + 37{,}900}{860}$$

$$= 487.8 \, Nm/kg$$

Since $W\eta_p$ also is the head produced, $\underline{\underline{H = 487.8 \, Nm/kg}}$ Answer

$$\dot{m} = (0.0025)(860) = 2.15 \, kg/s$$

Power input is thus

$$P = (2.15)(487.6)/0.6 = 1747 \, kJ/s = \underline{\underline{1.75 \, kW}}$$

Answer

It can be shown that $NSPH = (p_1 - \overset{\circ}{p}_v - F_s)/\rho g - Z_1$, where p_1 is the pressure at the surface of the tank, $\overset{\circ}{p}_v$ the vapor pressure of the fluid and F the friction in the suction line

$$NSPH = \frac{1}{9.806}\left[\frac{101{,}300 - 26{,}400 - 3450}{860}\right] - 1 = 8.47 - 1$$

$$= 7.47 \, m$$

The available NSPH is larger than the net positive head of 3m required. The pump should be suitable.

Answer

12-5

$\dot{m}_v = 0.0063\ m^3/s$ at 15°C

$N_s = 1800$ rpm

$\Delta p = 138$ kPa

For an ideal centrifugal pump, the speed of the impeller is proportional to the rate of fluid discharge. The head developed by an ideal pump is proportional to the square of the impeller speed. Thus from Eq. (12-17)

$$m_{v,2} = m_{v,1}\left(\frac{N_{r,2}}{N_{r,1}}\right)$$

$$= (0.0063)(1200/1800) = 0.0042\ m^3/s$$

$$\text{or } \dot{m}_2 = \dot{m}_{v,2}\,\rho = (0.0042)(998) = \underline{\underline{4.19\ kg/s}}$$

Answer

The new head with the reduction in impeller speed is given by

$$H = \left(\frac{N_{r_2}}{N_{r_1}}\right)^2\left(\frac{\Delta p}{\rho}\right)$$

$$= \left(\frac{1200}{1800}\right)^2\left(\frac{138{,}000}{998}\right) = 61.6\ Nm/kg$$

This is equivalent to $(61.6)/(9.806) = \underline{\underline{6.28\ m}}$

Answer

12-6 $\dot{m}_{H_2O} = 15.75$ kg/s, $\rho_L = 998$ kg/m³, $\mu_L = 1.120$ cP

Assume turbulent flow. From Eq. (12-15)

$$D_{i,opt} = 0.344\, \dot{m}_v^{0.45} \rho^{0.13} = 0.344(15.75/998)^{0.45}(998)^{0.13} = 0.130\text{m}.$$

Use 5 in nominal steel pipe, Schedule 40, D_i = 5.047 in = 0.128 m

From Fig. 12-3, $Re = 1280\, \rho\, \dot{m}_v / D_i \mu$

$$= (1280)(998)(15.75/998)/(0.128)(1.120) = 141{,}000$$

Flow is turbulent. To use Fig 12-20 to obtain cost of pump and motor need to develop equivalent pressure drop across pump.

$$\Delta p = (7.5 \text{ kNm/s})(998 \text{ kg/s})/((15.75 \text{ kg/s}) = 475.2 \text{ kPa}$$

$$(\text{m}^3/\text{s})(\text{kPa}) = (15.75/998)(475.2) = 7.5$$

From Fig 12-20, cost of pump and motor is $3600

From Fig. 12-4, cost of 0.128 m diameter carbon seamless pipe is $23/m

Pipe cost = (23)(305) = $7015

From Fig 14-4 (4th edition) the cost of fittings indered to Jan. 1, 2002 is $70 /tee

Fittings cost = (70)(40) = $2800

From Fig. 12-8, the cost of a gate valve, carbon steel, flanged, with a 860 rating, is $310/ valve

Valve cost = (310)(4) = $1240

From Fig. 12-12, the cost of 0.038 m 85% magnesia is $16 /m

Insulation cost = (16)(305) = $4880

Information from Table 6-5 indicates that the cost of installation of pumps is from 25 to 60% of the purchased equipment cost. Use an average value of 40%, or a factor of 1.4. Installed cost is then

Cost = (3600 + 7015 + 2800 + 1240 + 4880)(1.4) = $27,300

Answer

12-7

Two-step steam jet, suction pressure of 50 mm Hg

Vapor pressure of water at 15°C = 13.25 mm Hg

$$\% \text{ dry air} = \frac{(50-13.25)(29)}{(50-13.25)(29)+(13.25)(18)} = 0.817 \cong 82\%$$

Total weight of mixture/h = (10/0.817) = 12.25 kg/h

From Table 12-7 for 4.5 kg/h of mixture at 82% air, the steam requirement is

$$19 + (12/30)(4) = 20.6 \text{ kg/h}$$

Estimated total steam required for 12.25 kg/h is

$$(20.6)(12.25/4.5) = \underline{\underline{56 \text{ kg/h}}}$$

12-8

Compress methane at 25°C and 100 kPa to 6200 kPa

Determine the number of stages required. Try two stages:

$(6200/100)^{1/2} = 7.87$ this is too high, try three stages:

$(6200/100)^{1/3} = 3.95$ this is adequate. 3 stages required

The power required/stage is obtained with the use of Eq. (12-22b)

$$P = (2.78 \times 10^{-4})(0.085)(3600)(10^5)\left(\frac{1.31}{1.31-1}\right)\left[(3.95)^{0.2366}-1\right](10^{-3})/0.8$$
$$= 17.25 \text{ kW}$$

The power required by three stages is

$$p = (3)(17.25) = \underline{\underline{51.75 \text{ kW}}}$$

Answer

The exit temperature of the gas leaving the first stage is

$$T_2 = T_1 (p_2/p_1)^{(k-1)/k} = (298.1)(3.95)^{0.2366} = 412.6 \text{K} = \underline{\underline{139.5 °\text{C}}}$$

Answer

To obtain the cooling water required for the intercooler and aftercooler involves a heat balance. The heat load for one stage is

$$\dot{q} = \dot{m} C_p \Delta T = (0.085)(0.65)(2433)(139.5-25) = 15{,}400 \text{ J/s}$$

for three stages this totals to 46,200 J/s

Cooling water (assuming an average temperature of the water of 32°C) required is

$$\dot{m}_{H_2O} = \dot{q}/C_p \Delta T = 46{,}200/(4178)(15) = \underline{\underline{0.737 \text{ kg/s}}} \text{ or}$$

Answer

$\underline{\underline{7.4 \times 10^{-4} \text{ m}^3/\text{s}}}$

Answer

12-9

The use of a software package in the solution of Problem 12-8 provides results that are about 2% lower than those obtained for the power requirement by the hand calculation. The difference occurs essentially from the variation in the Cp/Cv ratio of the methane as it is compressed in each stage of compression since the process is not really adiabatic and reversible.

The software package calculation for the cooling water demand shows nearly a 6% deviation from the hand calculation. This deviation comes from the variation in the Cp/Cv ratio of the methane to obtain the exit gas temperature, the change in Cp of the methane affecting the heat load, and the use of the changing Cp value of water rather than using an average value of the water temperature. Obviously, the calculations with the computer software package are <u>more accurate.</u>

Answer

12-10

The mechanical energy balance, Eq.(2-2) in differential form is given as

$$dz + v\,dp + V\,dV = \int W - \int F \qquad \text{where } dz;\ V\,dV \text{ and } \int F = 0$$

$$W = \int v\,dp \qquad pv = RT/M$$

$$= \int_1^2 (RT/M)\,dp/dv = (RT/M)\ln p_2/p_1 = \frac{(8.314)(293)}{2.016}\ln\frac{4140}{1380}$$

$$= 1.327 \times 10^3\ \text{kJ/kg}$$

Actual power supplied is 224 kW

The kilograms of H_2 compressed is given by

$$(224\ \text{kJ/s})\left(\frac{0.55}{1.327\times10^3\ \text{kJ/kg}}\right) = \underline{\underline{0.093\ \text{kg/s}}}$$

Answer

For the isothermal and reversible case, the theoretical power is $p_1v_1\ln(p_2/p_1)$

$$\eta_{iso} = \frac{\text{theoretical power (isothermal)}}{\text{actual power}} = 0.55$$

For the adiabatic and reversible case, the theoretical power

$$\eta_{ad} = [k/(k-1)]\,p_1v_1\left[\left(p_2/p_1\right)^{(k-1)/k} - 1\right]$$

$$\eta_{ad} = \frac{\text{theoretical power (adiabatic)}}{\text{actual power}}$$

$$= \frac{\text{theoretical power (adiabatic)}}{\text{theoretical power (isothermal)}}(0.55)$$

$$= \frac{(1.4/0.4)\left[\left(\frac{4140}{1390}\right)^{0.4/1.4} - 1\right]}{\ln\left(\frac{4140}{1390}\right)}(0.55) = 0.646$$

$$= \underline{\underline{64.6\%}}$$

Answer

12-11

The actual suction density of the air is

$$\rho_1 = pM/RT = (10^5)(29)/(8314)(305) = 1.144\ \text{kg/m}^3$$

and the discharge density is

$$\rho_2 = (1.144)(103/101) = 1.178\ \text{kg/m}^3$$

Average density of the air removed is

$$\rho_{ave} = (1.144 + 1.178)/2 = 1.161\ \text{kg/m}^3$$

The mass flow rate is

$$\dot{m} = (4.7\ \text{std m}^3/\text{s})(1.144) = 5.38\ \text{kg/s}$$

In the mechanical energy balance, $Z_1 = Z_2$, $V_1 = 0$, $V_2 = 40\ \text{m/s}$, $F = 0$. The developed pressure is

$$(p_2 - p_1)\,\rho_{ave} = (103{,}000 - 100{,}000)/1.161 = 2585\ \text{Nm/kg}$$

The velocity head is

$$V_2^2/2\alpha = (40)^2/2 = 800\ \text{Nm/kg} \qquad \alpha = 1.0$$

$$W = [(p_2 - p_1)\,\rho_{ave} + V_2^2/2]/\eta$$
$$= (2585 + 800)(0.65) = 5210\ \text{Nm/kg}$$

The power requirement therefore is

$$P = (5210)(5.38) \cong 28{,}000\ \text{J/s} = \underline{\underline{28\ \text{kW}}}$$

Answer

12-12

The mixer in Example 12-6 has the following shape factors:

$S_1 = 1.0$, $S_2 = 1.0$ and $S_6 = 1.0$. S_3 and S_4 are also the same.

$\mu = 120\ \text{Pa}\cdot\text{s}$ $\rho = 1121\ \text{kg/m}^3$

The Reynolds number for this mixing system is

$$Re = D_a^2 N_r \rho/\mu = (0.6)^2(90/60)(121)/120 = 5.04 \quad \text{(laminar)}$$

For laminar flow, the Froude number may be neglected. Since the shape factors match those used to develop Fig. 12-40, this figure may be used to obtain the power requirement for this mixing problem. From Fig. 12-40, the power function is approximately 12.5.

$$P = \phi N_r^3 D_a^5 \rho$$

$$= (12.5)(90/60)^3(0.6)^5(1121)(10^{-3}) = \underline{\underline{3.67\ \text{kW}}}$$

There should be no diffence in the power requirement if baffles were added since with laminar flow the power function does not change as oberved in <u>Fig. 12-40.</u>

Answer

12-13

$D_t = 2\,m$ $D_a = 0.67\,m$ $C = 2\,m$ $\mu = 0.012\ Pa \cdot s$
$\rho = 1500\ kg/m^3$

Determine the Reynolds number for the mixing of the solution

$$Re = D_a^2 N_r \rho / \mu = (0.67)^2 (90/60)(1500)/0.012$$
$$= 84{,}200 \quad \text{(turbulent flow)}$$

For turbulent flow, $N_{P_0} = K_T$. Values for K_T are given in Table 12-9 for various types of impellers operating in tanks with four baffles ($S_5 = 0.1$)

$$K_T = 6.30$$

From Eq. (12-59)

$$P = K_T N_r^3 D_a^5 \rho$$
$$= (6.30)(90/60)^3 (0.67)^5 (1500) = 4306\ J/s = \underline{\underline{4.3\ kW}}$$

Answer

12-14

$D_i = 9\text{ m}$ $\quad p_i = 310\text{ kPa}$ $\quad T_i = 27°\text{C}$ no corrosion considered

Table 12-10 provides the relationship to determine the recommended wall thickness for a spherical tank under pressure as

$$t = \frac{Pr_i}{2SE_j - 0.2P} + C_c$$

where S is the maximum allowable stress and P the maximum allowable internal gauge pressure given in Table 12-10 for carbon steel

Assume lap weld, spot examined, $E_j = 0.80$

$$t = \frac{(209)(4.5)}{2(94{,}500)(0.8) - 0.2(209)} = \underline{\underline{0.0062\text{ m} = 0.244\text{ in.}}}$$

With the small wall thickness relative to r_i, it is satisfactory to obtain the volume occupied by the wall by multiplying the inside area with the minimum wall thickness.

$A_{sphere} = 4\pi r_i^2$ $\qquad V_{sphere} = (4/3)\pi r_i^3$

$\rho_{steel} = 7826\text{ kg/m}^3$ (from Table D-10)

Cost of steel used in the tank based on $1.10/kg is

$$\text{Cost} = 4\pi(4.5)^2(0.0062)(7826)(1.10 = \$13{,}600$$

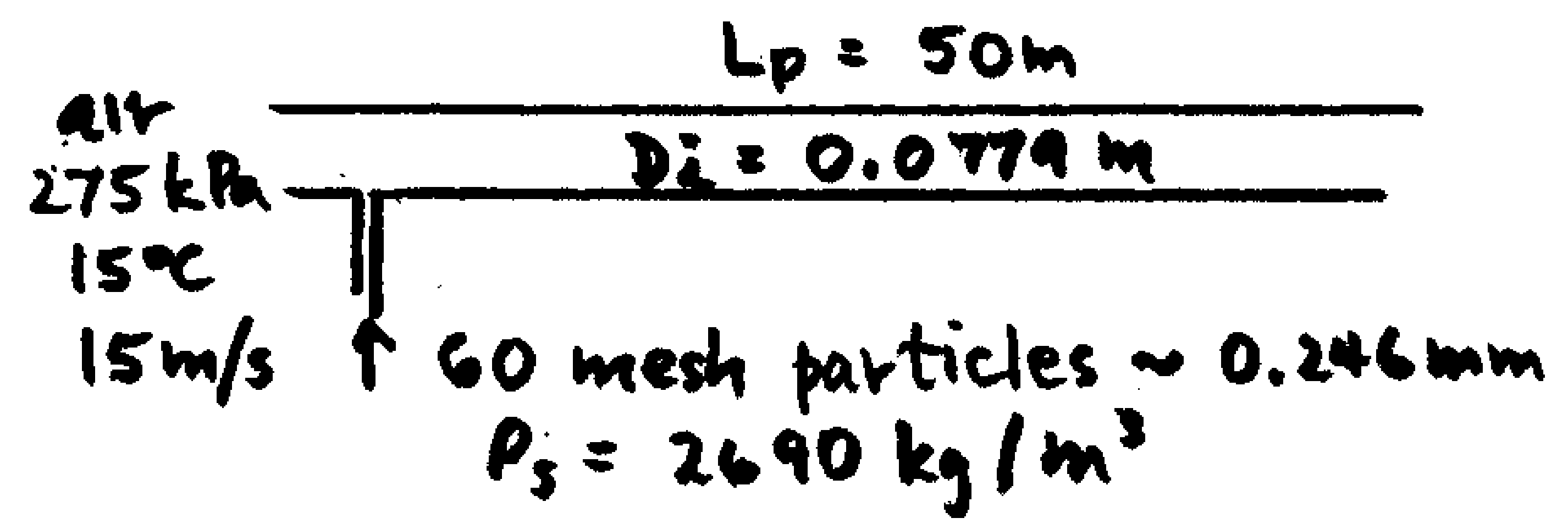
Cost of tank based on Fig. 12-52 is about $205,000 for a tank capacity of 381 m^3

Fraction of tank cost due to the steel required is

$$13{,}600/205{,}000 = 0.066 = \underline{6.6\%}$$

Answer

12-15



Determine Δp
$m_p/m_{air} = 4$

Density of air at the inlet is

$\rho_{air} = (275)(29)/(8.314)(288) = 3.33 \text{ kg/m}^3$

$G = \rho V = (3.33)(15) = 49.9 \text{ kg/m}^2\text{·s}$

$Re = DG/\mu = (0.0779)(49.9)/(1.8\times10^{-5}) = 2.16\times10^5$

From Fig. 12-1, f = 0.0045 for commercial pipe
Assume that the average density of the air in the flow process is 3.30 kg/m³. Calculate the density of the air at the discharge to verify ρ_{ave} for pure air flow from

$\Delta p = p_1 - p_2 = (G^2/\rho_{ave})(2fL_p/D + (2.303/\alpha)\log(p_1/p_2)$

neglect the log term to simplify the calculation

$= (49.9)^2(2)(0.0045)(50)/(3.30)(0.0779) = 4359$ Pa

$p_2 = p_1 - \Delta p = 275{,}000 - 4359 = 270{,}641$ Pa

$\rho_2 = (270.6)(29)/(8.314)(288) = 3.27 \text{ kg/m}^3$

Thus, $\rho_{ave} = (3.33 + 3.27)/2 = 3.30 \text{ kg/m}^3$ ρ_{ave} verified

Energy is required to move the solid particles and this is obtained from the air through the action of drag forces between the air and the solid. This is represented by

$E = r\,[(p_2-p_1)/\rho_s + (V_2^2-V_1^2)_s/2 + g(z_2-z_1)]$ where r is the solid/gas ratio

Assuming that the pressure drop is a small fraction of the absolute pressure, the air can be considered as an incompressible fluid of constant average density. Neglecting change in velocity head and assuming the kinetic energy factor to be unity, the mechanical energy balance provides the required energy. This is given by

$(p_2-p_1)/\rho_{ave} + g(z_2-z_1)_{air} = -E + \Sigma F$ (1)

where ΣF is the total friction in the stream
Eliminating E from both equations and solving for $p_1 - p_2$ gives

$$p_1 - p_2 = \frac{g(1+r)(z_2 - z_1) + (r/2)(V_2^2 - V_1^2)_s + \Sigma F}{1/\rho_{ave,air} + r/\rho_s}$$

The total friction term is the product of β and that from the flow of the air by itself and is given by

$\Sigma F = 2\beta f L V_{avg}^2 / g D$ where V_{ave} is the average superficial velocity of the air in the pipe

An empirical relation for the relative pressure of the solid is given by Vogt and White (<u>Ind</u>. <u>Eng</u>. <u>Chem</u>. 40:1731(1948)).

$$\beta = a(D/D_p)^2(\rho_{ave} r/\rho_s Re)^k + 1$$

According to Vogt and White, $a = 8$ and $k = 0.57$ for the conditions in this problem. This permits evaluation of the pressure drop for the combined air and solid flow.

$D_p = 60$ mesh $= 0.000246$ m

$$\beta = 8.0\left(\frac{0.0779}{2.46 \times 10^{-4}}\right)^2 \left(\frac{(3.30)(4)}{(2690)(2.16 \times 10^5)}\right)^{0.57} + 1 = 36.3$$

Substituting this into Eq. (1), noting that $z_1 = z_2$ and the velocity of the air and solid at the discharge is 15 m/s

$$p_1 - p_2 = \frac{(4/2)[(15)^2 - 0] + 2(36.3)(0.0045)(50)(15.12)^2/(9.806)(.0779)}{(1/3.30) + (4/2690)}$$

$\Delta p = 17{,}520$ Pa $= \underline{\underline{17.5 \text{ kPa}}}$ (6.4% of original press.)

Answer

12-16

Equation (12-60) can be used to obtain the power consumption under the new conditions. Since the only variables being considered are the feed rate and the volume-surface mean diameter, Eq. (12-60) can be used for the original operation and then for the new operation. It will not be necessary to convert units of Eq. (12-60) since the conversion factors can be included in an overall constant.

For the original operation

$$\frac{P_1}{\dot{m}_s} = C\left(\frac{1}{D_{pp}^{1/2}} - \frac{1}{D_{pf}^{1/2}}\right)$$

where C includes the work index as well as the unit conversions which remain the same for both conditions

$$\frac{6.35}{10{,}000} = C\left[\frac{1}{(0.005)^{1/2}} - \frac{1}{(0.069)^{1/2}}\right] = 10.335\,C$$

For the new operation

$$\frac{P_2}{9{,}000} = C\left[\frac{1}{(0.004)^{1/2}} - \frac{1}{(0.069)^{1/2}}\right] = 12.0\,C$$

Dividing the second equation by the first, permits evaluation of P_2

$$\frac{P_2\,(10{,}000)}{(6.35)(9{,}000)} = \frac{12.0}{10.335}$$

$$P_2 = (12)(6.35)(9{,}000)/((10.335)(10{,}000)) = \underline{\underline{6.64\ \text{kW}}}$$

Answer

Chapter 13

13-1. (a) $$\frac{V_R}{F_{io}} = \frac{X}{-r_i}$$

8% drop in conversion

$$X_{new} = .92\, X_{orig.}$$

$$\Longrightarrow \underline{V_{R_{new}} = .92\, V_{R_{orig}}}$$

(b) Yes, it is possible that the decrease is due to the malfunctioning of one mixer; however, it is not possible to tell which one since no matter which one fails, the process goes from 4 reactors in series to 3.

13-2 (a) space time $= \frac{V_R \, c_{iF}}{F_i}$

$$\frac{V_R}{F_i} = \frac{X}{-r_i} \quad ; \quad X = 0.5$$

$$-r = k\,[CH_3OH]^2[(C_6H_5)_3CCl]$$

$$k = 4.48\times10^{-3}\left(\frac{m^3}{kgmol}\right)^2\Big/ s \quad @\ 298K\ ;\ 101kPa$$

feed conc: $0.11\ \frac{kg\,mol}{m^3}$ triphenyl methyl chloride

$0.055\ \frac{kgmol}{m^3}$ methanol

$$\frac{X}{-r_i} = \frac{0.5}{4.48\times10^{-3}\left(\frac{m^3}{kgmol}\right)^2\Big/s\left[\frac{.055}{2}\right]^2\left[.11-\frac{.055}{2}\right]\left(\frac{kg\,mol}{m^3}\right)^3}$$

$$= 1.789\times10^6\ \frac{m^3\cdot s}{kg\,mol}$$

$$\frac{V_R}{F_i}c_{iF} = 1.789\times10^6\ \frac{m^3\cdot s}{kg\,mol}\cdot 0.055\ \frac{kgmol}{m^3} = \underline{\underline{9.839\times10^4\ s}}$$

(b) order of reaction is positive for both reactants ⇒ minimize volume with PFR or batch

13-3. (a) $\underset{(1)}{BrCN} + \underset{(2)}{CH_3NH_2} \longrightarrow NH_2CN + CH_3Br$

$$\frac{V_R}{F_i} = \frac{X_i}{-r_i}$$

$V_R = 2\ m^3$

$F_1 = .01\ \frac{kgmol}{s}$; $F_2 = .02\ \frac{kgmol}{s}$

For (1) $$\frac{V_R}{F_1} = \frac{\frac{C_{o_1} - C_1}{C_{o_1}}}{k\, C_1 C_2}$$

$k = 2.22\ \frac{m^3}{kgmol \cdot s}$

C_o - is a combined feed conc.

$C_{o_1} = .05\ \frac{kgmol}{m^3}$ $\qquad C_{o_2} = .1\ \frac{kgmol}{m^3}$

Solve for C_1 $\qquad (C_2 = C_1 + .05)$

$C_1 = 0.0196\ \frac{kgmol}{m^3}$

$$X_1 = \frac{.05 - .0196}{.05}$$

$$\underline{\underline{X_1 = 0.607}}$$

Similarly, for C_2

$$\frac{V_R}{F_2} = \frac{\frac{C_{o_2} - C_2}{C_{o_2}}}{k\, C_1 C_2}$$

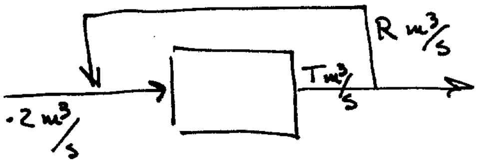
$$X_2 = \frac{C_{o_2} - C_2}{C_{o_2}} = \frac{.1 - .0696}{.1}$$

$$= \underline{\underline{0.304}}$$

(b) recycle of 90% of unreacted reactants
20% of water
20% of product

assume product recycle will not dilute reactants.

recycle flowrate:



$T = .2 + R$

$R = .2T$

$T = 1.25(.2\ m^3/s) = .25\ m^3/s$

$R = .05\ m^3/s$

New feed rates:

(1) $.0196 \frac{kg\,mole}{m^3} \cdot \frac{.2\,m^3}{sec} = .00392 \frac{kg\,mole}{s}$ effluent

90% of this is $.003528 \frac{kg\,mol}{s}$

$F_{1_{new}} = .01 \frac{kg\,mol}{s} + .003528 \frac{kg\,mol}{s}$

$F_{1\,new} = .013528\ mol/sec$

$C_{o_1} = \frac{.013528\ mol/s}{.25\,m^3/s} = .05411\ kg\,mol/m^3$

(2) $.0696 \frac{kg\,mole}{m^3} \cdot \frac{.2\,m^3}{s} = .01392 \frac{kg\,mole}{s}$ effluent

90% → $.012528 \frac{kg\,mol}{s}$

$F_{2_{new}} = .02 + .012528 = .032528 \frac{kg\,mol}{s}$

$C_{o_2} = \frac{.032528 \frac{kg\,mol}{s}}{.25\,m^3/s} = 0.1301 \frac{kg\,mol}{m^3}$

$c_2 = c_1 + .076$

(1) $$\frac{V_R}{F_1} = \frac{\frac{C_{o_1} - C_1}{C_{o_1}}}{k\, C_1 (C_1 + .076)}$$

solve for C_1

$$C_1 = 0.0200 \ \frac{\text{kg mol}}{\text{m}^3}$$

$$X_1 = \frac{.05411 - .0200}{.05411}$$

$$= \underline{\underline{0.63}}$$

(2) $$X_2 = \underline{\underline{0.26}}$$

13-4

(1) $C_3H_6 + \frac{3}{2} O_2 \rightarrow C_3H_4O_2 + H_2O$

(2) $C_3H_6 + \frac{5}{2} O_2 \rightarrow C_2H_4O_2 + CO_2 + H_2O$

(3) $C_3H_6 + \frac{9}{2} O_2 \rightarrow 3CO_2 + 3H_2O$

$$-r_i = A_i \exp\left(\frac{-E_A}{RT}\right) P_{prop} P_{O_2}$$

Reaction	E_A, kJ/kgmol	A_i (kg mol/m³)(kPa)$^{-2}$(s)$^{-1}$
1	3590	4.42×10^1
2	5980	5.08×10^4
3	4780	2.45×10^2

Want to maximize r_1 relative to r_2 & r_3

$$\frac{r_1}{r_2} = \frac{A_1}{A_2} \exp\left[\frac{E_{A_2} - E_{A_1}}{RT}\right]$$

$$\frac{r_1}{r_3} = \frac{A_1}{A_3} \exp\left[\frac{E_{A_3} - E_{A_1}}{RT}\right]$$

since $E_{A_2}, E_{A_3} > E_{A_1}$, Low T's maximize ratio's & hence selectivity for rxn 1.

All 3 rxn's are 1st order in reactants, so altering concentrations will not change selectivity, but equal amounts of propylene & O_2 are required. Pressure changes will not affect selectivity.

13-5.

$$-r = kc^2$$

$$k = 5.4\times10^7 \exp\left(\frac{-100{,}250}{RT}\right) \quad m^3/kgmol\cdot s$$

3 CSTR's

$V_1 = 8\ m^3 \quad V_2 = 6\ m^3 \quad V_3 = 4\ m^3$

$T = 500K$

$P = 101\ kPa$

$v = 7\ m^3/day$

(a) assume ideal gas

$$C_{AO} = .02437\ \frac{kgmol}{m^3}$$

$$\varepsilon = -\frac{1}{2}$$

$$X_A = \frac{1-\frac{C_A}{C_{AO}}}{1-\frac{C_A}{2C_{AO}}}$$

$$-r = kC^2 = \frac{v\,C_{AO}X_A}{V}$$

solve for X_A, C_A for each reactor, remembering to adjust v.

Reactor	X_A	C_A
1	.742	.00999
2	.224	.00873
3	.091	.008315

overall

$$\boxed{X_A = 0.794}$$

(b) batch reactor - 50% time

(see Levenspiel)

2nd order batch

$$\frac{(1+\varepsilon_A)X_A}{1-X_A} + \varepsilon_A \ln(1-X_A) = k C_{A0} t$$

$\varepsilon_A = -\frac{1}{2}$; $X_A = 0.794$

$k = 1.817\times10^{-3} \frac{m^3}{kgmol\, s}$

$C_{A0} = .02437 \frac{kg\,mol}{m^3}$

$\Rightarrow t = 61359\ s$

$v = 8.1\times10^{-5} \frac{m^3}{s}$

to process that quantity & accounting for operating 50% of the time

$V = 2vt$

$$\boxed{V = 9.94\ m^3}$$

13-6 Slurry tank reactor

$2.6 \frac{kgmol}{m^3}$ glucose $10^{-3} \frac{m^3}{s}$

120% H_2 $T = 423 K$

$-r_{glucose} = k\, c_{H_2}^{0.6}\, c_{glu}$ $k = 3.76 \times 10^{-4} \left(\frac{kgmol}{m^3}\right)^{-.6} / s$

$P = 10{,}000 \text{ kPa}$ $S_{H_2} = 5 \frac{kgmol}{m^3}$

$V = 2 m^3$ CSTR

$\frac{V}{F} = \frac{X_A}{-r_A}$ $F = \frac{2.6 kgmol}{m^3} \cdot 10^{-3} \frac{m^3}{s}$

$= 2.6 \times 10^{-3} \frac{kgmol}{s}$

$X_A = \frac{2.6 - C_{glu}}{2.6}$

$-r_A = k \left(5 \frac{kgmol}{s}\right)^{.6} C_{glu}$

Solve for C_{glu}

$C_{glu} = 0.874$

$\boxed{X_A = 0.66}$

13-7 $-r_{PO} = k C_{PO}$

$$k = 4.71 \times 10^9 \exp\left(\frac{-63{,}010}{RT}\right)/s$$

$T = 300\ K$

Feed : 10 wt % PO (aqueous) .01 m^3/s

0.1 wt % H_2SO_4 (catalyst) .01 m^3/s

90% conversion

$$\frac{V}{v} = -\frac{1}{k} \ln \frac{C_A}{C_{A0}}$$

$$k = 0.0503/s$$

$$\frac{C_A}{C_{A0}} = .1$$

$$v = .02\ m^3/s$$

$$\boxed{V = 0.9155\ m^3}$$

PROBLEM 13-8

Correction: Add, the ethanol yield is 0.47 kg ethanol/kg of glucose consumed. The correct unit for the rate coefficient, *k*, is 1/s, not kg/m^3·s

a) The time for a batch reaction is given by Eq. 13-8. Since N_i/V_R is the reactant concentration, glucose in this case, the equation can be written as

$$\theta = \int_{c_{\text{glucose},i}}^{c_{\text{glucose},f}} d(c_{\text{glucose}})/(r_{\text{glucose}}) \approx \sum_{j=1}^{j=g} (\Delta c_{\text{glucose}})/(r_{\text{glucose}})_{j,\text{average}}$$

where $c_{\text{glucose},f}$ is the final glucose concentration = (0.05*10) kg/m^3, $c_{\text{glucose,i}}$ the initial glucose concentration=10 kg/m^3, $\Delta c_{\text{glucose}}$ the incremental change of glucose concentration in the numerical integration (a negative quantity), and $g=(c_{\text{glucose},f} - c_{\text{glucose},i})/\Delta c_{\text{glucose}}$. Because the rate is a complicated form in concentration, a numerical integration is recommended. Any acceptable numerical integration routine may be used. Here, integration was performed with a spreadsheet using the trapezoidal approximation. The Δ-increment size was decreased in successive integrations until the result does not change significantly with increment size. The term $(r_{\text{glucose}})_{j,\text{average}} = (r_{\text{glucose},j-1} + r_{\text{glucose},j})/2$ for integration by the trapezoidal method.

The reaction rate is given by

$$r_{\text{glucose}} = -(1.53*10^{-3})\Big(1-(c_{\text{ethanol}}/93)\Big)^{0.52}\Big((c_{\text{glucose}})(c_{S.cerevisiae})/(c_{\text{glucose}}+1.7)\Big)$$

The concentrations are related as follows:

$$c_{\text{glucose},j} = c_{\text{glucose},i} + \sum_{j=1}^{j} \Delta c_{\text{glucose}} \qquad c_{\text{ethanol},j} = -(0.47)\sum_{j=1}^{j} \Delta c_{\text{glucose},j}$$

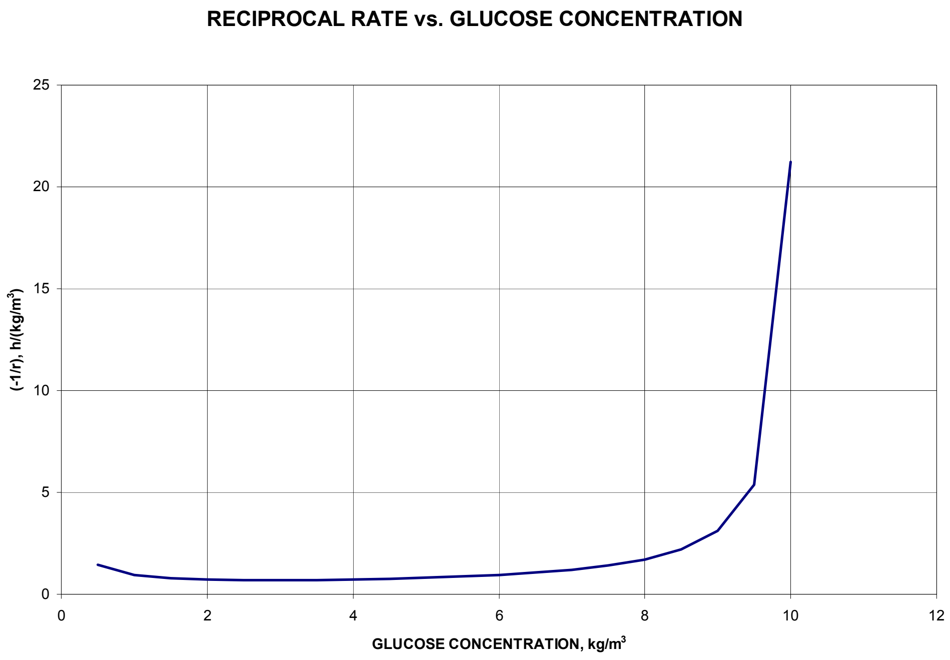
$$c_{S.cereisiaei,j} = 0.01 - (0.06)\sum_{j=1}^{j} \Delta c_{\text{glucose}}$$

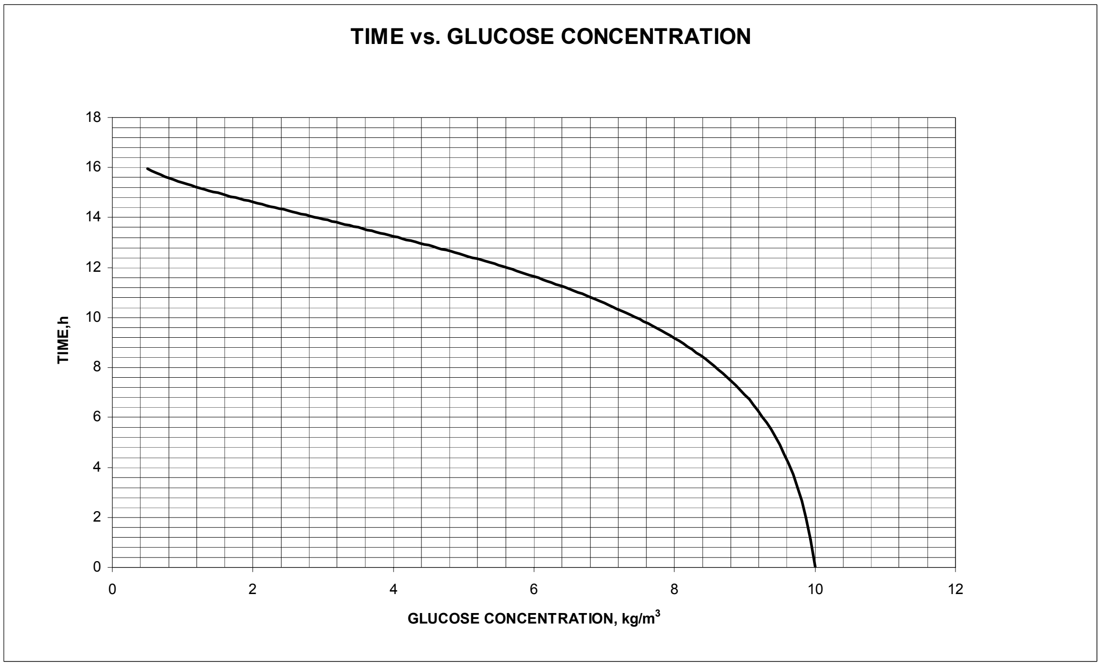
The reciprocal of rate as a function of c_{glucose} and the value of the integral (which is time) as a function of c_{glucose} are shown on the next page.

The result for these conditions is

<u>θ = 16.0 h</u> <u>ANSWER</u>

PROBLEM 13-8 (continued-1)





PROBLEM 13-8 (continued-2)

b) The PFR will have a volume of 7 m^3. Eq. (13-13a) can be written in terms of the total volumetric flowrate V, assumed constant, and glucose concentration as

$$V_R = (V)\int_{c_{\text{glucose},i}}^{c_{\text{glucose},f}} d(c_{\text{glucose}})/(r_{\text{glucose}})$$

The integral must be reevaluated with the new initial concentration of $c_{S.cerevisiae}$ (=0.9 kg/m^3). The result is a value of 2.23 h.

Thus, the flow rate needed is

$V = V_R/(2.23) = (7\ m^3)/(2.23\ h) = \underline{\underline{3.1\ m^3/h}}$ ANSWER

c) The volume of a single CSTR for this reaction, assuming that the feed concentration of $c_{S.cerevisiae}$ the same as in part b), is obtained from Eq. (13-18) in the volumetric flowrate/concentration form as

$$V_R = V(c_{\text{glucose},f} - c_{\text{glucose},i})/(r_{\text{glucose},f})$$

where $r_{\text{glucose},f}$ is the rate of glucose reaction at the conditions leaving the reactor.

$$V_R = (3.1\ m^3/h)(0.5 - 10\ kg/m^3)/(-1.79\ kg/m^3\cdot h) = \underline{\underline{16.6\ m^3}}\ \text{ANSWER}$$

From Fig. 13-15, choosing a glass-lined, jacketed, agitated reactor, the cost is estimated to be

$ 120,000 ANSWER

For the case of three equal volume reactors, an iterative solution is required, wherein reactor sizes are represented as rectangular areas on a reciprocal rate vs. concentration graph. The outlet concentration is varied until the three reactor volumes (rectangular areas on the graph) are equal. See Ex. 13-7. The result, as shown on the graph on the following page, is that the value of the integral for each of the three reactors is 0.90 h. Thus, the volume of each reactor is

$V_R = V*(0.90) = (3.1)(.90) = 2.8\ m^3$

and the total volume of the 3 reactors = 3*2.8 = $\underline{\underline{8.4\ m^3}}$ ANSWER

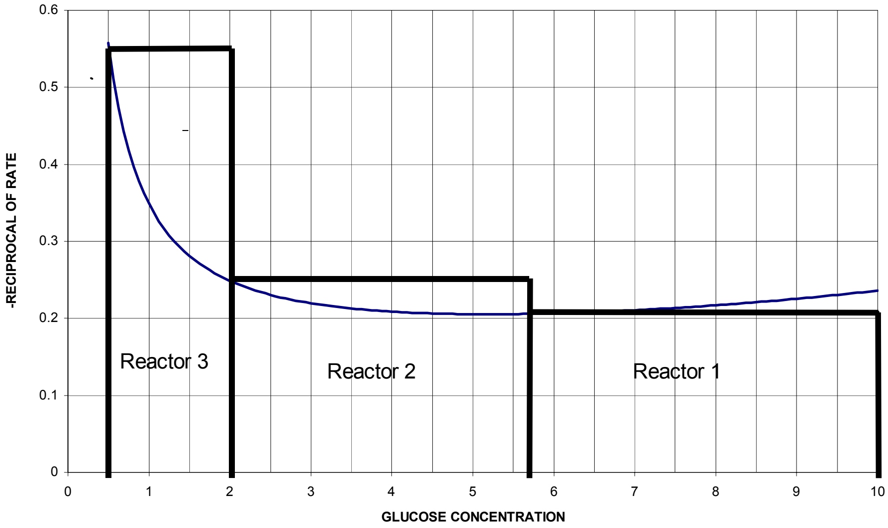
From Fig. 13-15, again choosing the glass-lined reactor, the cost is estimated to be

$ 50,000 each, or $150,000 for the three ANSWER

Even though the total volume is just half of that for a single CSTR, the cost is greater for the three reactors. This is a consequence of the economy of scale for such reactors. There are other factors, such as flexibility of multiple reactors that would influence such a selection. If a material other than glass-lined was chosen, the cost figures would be different, but the relative values would be very similar.

PROBLEM 13-8 (continued-3)

RECIPROCAL RATE vs. GLUCOSE CONCENTRATION



PROBLEM 13-9

This solution uses the reciprocal rate vs. concentration curve developed in part b) of problem 13-8 and shown below. The term "best arrangement" is used to mean the combination of the given reactors that gives the greatest degree of conversion for the given reactor sizes and types, and the flowrate and conditions of part b), problem 13-8.

Each of the reactor types and volumes suggested in parts a) through d), will be matched with the reciprocal rate vs. concentration curve to find the lowest concentration achievable.

This reaction is autocatalyzed, and the reaction rate initially increases as the glucose concentration decreases, because the biomass concentration increases. Eventually, the glucose concentration decreases sufficiently to cause the reaction rate to decrease; the increase of ethanol concentration contributes somewhat to the decrease also. Two guidelines are helpful. For PFR combinations, the order or sequence of PFR reactors makes no difference, because the reactor volume is proportional to the area under the reciprocal rate vs. concentration curve. So this area and the resulting outlet concentration are fixed, regardless of the order. The order of CSTRs is important, however, because the conditions at the reactor outlet determine the reaction rate in the reactor. Starting at the initial glucose concentration, as long as the reciprocal rate is decreasing (rate is increasing), the conversion achieved with a given volume of CSTRs will be greater than that for the same volume of PFRs, and the larger CSTR should be be used first. These guidelines are illustrated by the following solution.

The following solution is based on the conditions for part b) of problem 13-8. A graph of reciprocal rate vs. concentration is shown in the solution for problem 13-8, part b), and the value of the integral vs. concentration from that problem is shown on the following page.

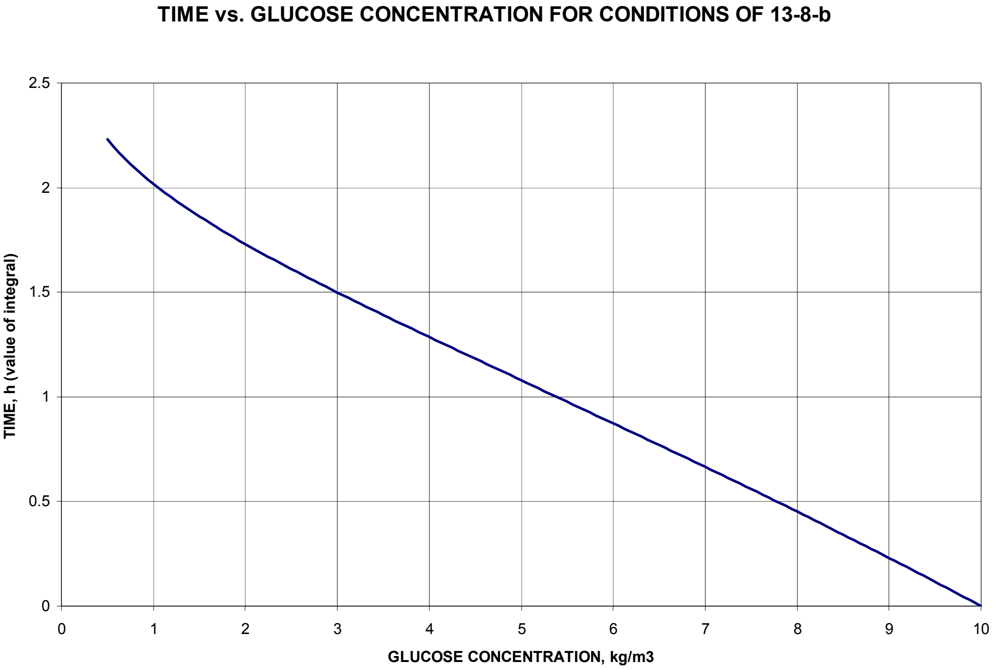
a) Eq. (13-13a) for PFR, in concentration terms, is

$$V_R = (V)\int_{c_{\text{glucose},i}}^{c_{\text{glucose},f}} \mathrm{d}(c_{\text{glucose}})/(r_{\text{glucose}});\ \text{thus, Intergal} = V_R/\mathrm{V} = (2+1)/3.1 = 0.968$$

For two PFRs in series in either order, the value of the integral is fixed, so the result is independent of the arrangement. The only variable left is $c_{\text{glucose},f}$ and this is read from the abcissa of the graph on the following page for the ordinate equal to 0.968.

The result is, $\underline{\underline{c_{\text{glucose},f} = 5.6\ \text{kg/m}^3}}$ ANSWER

PROBLEM 13-9 (continued-1)



PROBLEM 13-9 (continued-2)

b) For CSTRs, Eq. 13-18 in concentration form, gives the volume is by

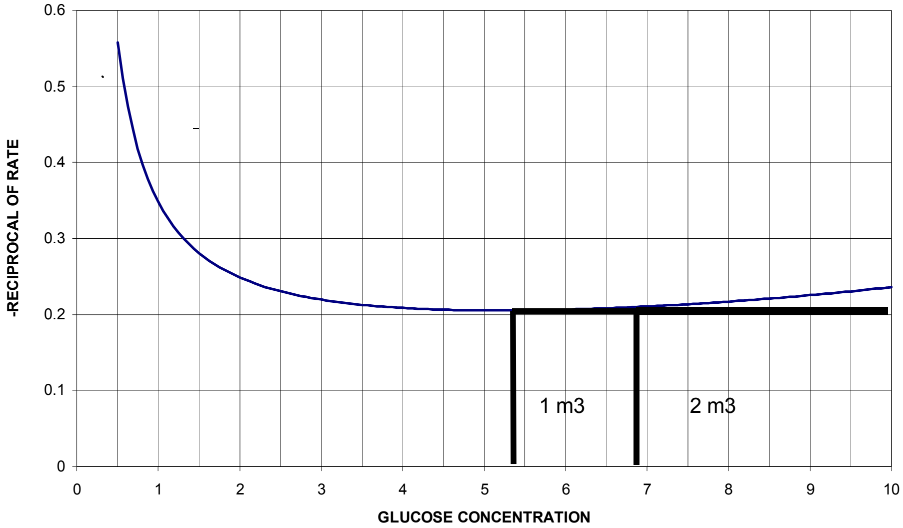
$$V_R = V(c_{\text{glucose},f} - c_{\text{glucose},i})/(r_{\text{glucose},f})$$

A value of $c_{\text{glucose},f}$ is assumed, the corresponding $(1/r_{\text{glucose},f})$ is found from the graph below, and iteration is continued until the foregoing equation is satisfied. Since sequence matters for the two unequal CSTR case, both combinations must be tried to find the one that gives the lowest $c_{\text{glucose},f}$. The better combination, the 2-m^3 CSTR followed by the 1-m^3, is shown on the graph. In this case, the first reactor reduces the glucose concentration to 6.9 kg/m^3 and the second to 5.3 kg/m^3. So the final result is, a bit lower than the value obtained with the 2 PFRs in part a).

$c_{\text{glucose},f} = 5.3$ kg/m^3 ANSWER

For comparison, if the 1-m^3 is placed first, $c_{\text{glucose},f} = 5.5$ kg/m^3, so the result is better with the bigger reactor first.

RECIPROCAL RATE vs. GLUCOSE CONCENTRATION



c) If the choice is between the four reactors given, it is obvious that the 2-m^3 PFR and the 2-m^3 CSTR should be used. From the above graph and the results in parts a) and b), it is clear that putting the 2-m^3 CSTR first, followed by the the 2-m^3 PFR will achieve the lowest value of $c_{\text{glucose},f}$. The value is obtained by starting with the 6.9 kg/m^3 from the 2-m^3 CSTR, finding the time on the first graph corresponding to that concentration (0.7h), and then adding to that the space time of the 2-m^3 PFR (0.65 h) and then, from the same graph, finding the concentration corresponding to the total time (1.35 h). The result is

$$\underline{\underline{c_{\text{glucose},f} = 3.7 \text{ kg/m}^3 \quad \text{ANSWER}}}$$

Note that some might interpret the statement of this part to mean the use of one PFR and one CSTR of equal, but unspecified, volumes, to reach the original goal of 95 % conversion. The result for this case is a volume of 3.3 m^3 for each reactor.

d) If all four reactors available reactors are combined, it can be deduced from the results above that it is best to put the 2-m^3 CSTR first, the 1-m^3 CSTR second and to follow these with the 2 PFRs in either order. The result (obtained as in part c) is

$$\underline{\underline{c_{\text{glucose},f} = 1.2 \text{ kg/m}^3 \quad \text{ANSWER}}}$$

e) The glucose conversion = $[(c_{\text{glucose},i} - c_{\text{glucose},f})/\ c_{\text{glucose},i}](100)$, and the values are as follows:

a) 44%
b) 47%
c) 63%
d) 88%

PROBLEM 13-10

Corrections: The heat of reaction is 89.98 MJ/kg mol CH_3OH, not kJ/kg mol. The symbol $\bar{p}_i$ represents fugacity, in kPa, of component *i*, which equals the partial pressure, $p_i = Py_i$, in kPa, when the fugacity coefficient is 1.

a) At equilbrium, the numerator of the reaction rate expression is equal to zero, or,

$$K_{eq} = (p_{CH3OH})/(p_{CO})(p_{H2})^2 = P(y_{CH3OH})/(P^3(y_{CO})(y_{H2})^2)$$

The mol fractions can be expressed in terms of the conversion of CO, *X*, starting with 1 mol of CO and 2 moles of hydrogen, as follows:

$$CO + 2H_2 \rightleftharpoons CH_3OH$$

$(1\text{-}X)$ $(2\text{-}2X)$ (X) (moles). At any conversion, the total number of mols is the sum and equals $(3\text{-}2X)$. So the mole fractions are given by,

$$y_{CO} = (1\text{-}X)/(3\text{-}2X); \quad y_{H2} = (2\text{-}2X)/(3\text{-}2X); \text{ and } y_{CH3OH} = X/(3\text{-}2X)$$

Substituting these terms into the equilibrium expression gives,

$$K_{eq} = [X/(3\text{-}2x)]/[P^2(4(1\text{-}X)^3/(3\text{-}2X)^3)], \text{ or } 4K_{eq}P^2 = X(3\text{-}2X)^2/(1\text{-}X)^3$$

This latter equation can be solved by entering the known values and iterating on *X* to find the conversion at equilibrium.

The result is $X_{eq} = 0.610$ ANSWER

b) A packed-bed reactor is specified, and is best modeled as a plug-flow reactor. The catalyst-to-feed ratio is given by Eq. 13-15. In order to find the ratio, the integral of $(dX_{iz})/(-r_{ic})$ must be evaluated between the limits $X_{io} = 0$ and $X_{iz} = 90$ percent of the equilibrium value from part a (0.9*0.61=0.549).

The reaction rate for CO is written as

$$-r_{CO} = \frac{\{[4P^3(1 - X)^3/(3\text{-}2X)^3] - (P)(X)/[(3 - 2X)(3.18*10^{-7})]\}}{[23400 + 126(P)(1\text{-}X)/(3\text{-}2X) + (47)(P)(2\text{-}2X)/(3\text{-}2X)]^2}$$

Because the rate rate expression is a complex function of *X*, analytical integration is impractical and a numerical evalution is easier. Any numerical integration method can be used. Here, the value of $-r_{CO}$ has been calculated for values of *X* incremented from 0 by a constant until $X = 0.549$ has been passed. The area under the curve of $1/(-r_{CO})$ vs. *X* was obtained by the trapezoid rule, and the calulation was repeated with a smaller increment size until the value of the integral does not change significantly. Graphs of $1/(-r_{CO})$ vs. *X* and the value of the integral vs. *X* are shown on the second following page. Interpolating for $X = 0.549$ gives a value for the integral equal to 7.2.

By Eq. 13-15, this is the value of W_c/F_{io}, i.e.

W_c/F_{io} = 7.2 kg catalyst per kg mol CO/min ANSWER

PROBLEM 13-10 (continued-1)

c) For a methanol production rate of 50 metric tons/h, the CO feed rate to the reactor must be

(50,000 kg CH_3OH/h)(1 kg mol/32 kg)(1 kg mol CO/0.549 kg mol CH_3OH)(1h/60 min) = 47.43 kg mol CO/min. And therefore the reactor volume (filled with catalyst) is

(7.2 kg cat./kg mol CO/min)(47.43 kg mol CO/min)/(653 kg cat./m^3 reactor vol.) = <u>0.523 m^3 reactor volume</u> <u>ANSWER</u>

Because the large heat of reaction must be removed, it is expected that a shell-and-tube reactor will be needed, with the catalyst packed inside the tubes.

The heat duty for the reactor is

(0.43 kg mol CH_3OH/s)(89,980 kJ/kg mol) = 38,650 kJ/s

The tube number, length and diameter must be selected so as to provide sufficient volume for the reaction and sufficient area for the heat transfer. This is done as illustrated in Ex. 13-8. Using 0.0508-m diameter and 6.09-m tubes from the example, the number of tubes needed is

$(0.523\ m^3)/(0.00956\ m^3 \text{ per tube}) = 55$

The available surface area is then,

$(0.973\ m^2/\text{tube})(55 \text{ tubes}) = 53.5\ m^2$

The heat transfer capability must be checked. Assume a $\Delta T_{\text{log mean}}$ of 50ºC and a heat transfer coefficient of 0.25 kW/$m^2 \cdot K$. The heat transfer rate is then

$Q = (0.25)(53.5)(50) = 2140$ kW= 669 kJ/s

Clearly in this case, heat transfer dominates the sizing of the reactor. The area, A, required to handle the heat removal is

$A = (38,650)/(.25)(50) = 3090\ m^2$

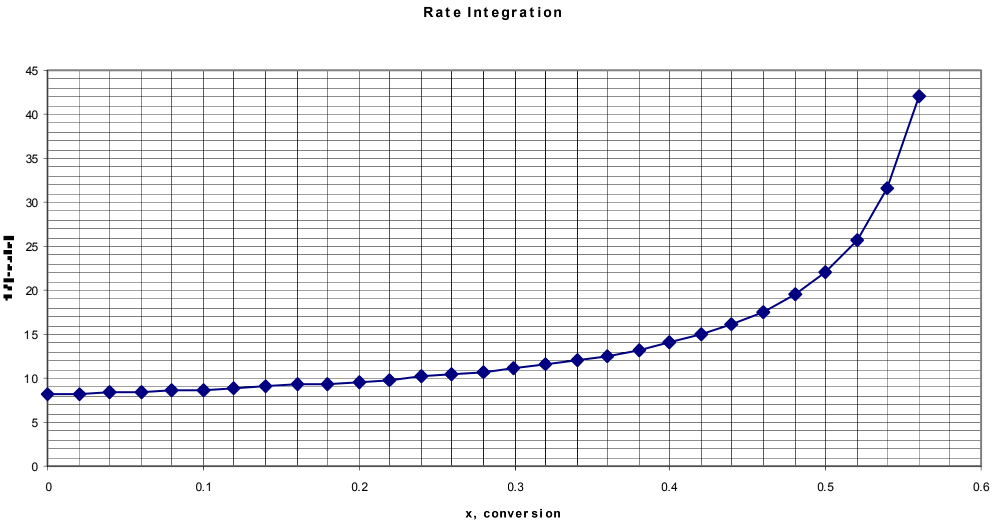
and the number of tubes required is 3090/(0.973) = 3176. From Fig. 14-17, it is found that the maximum size for a standard floating head shell-and-tube exchanger is 1000-m^2 area, with a base cost of $70,000. Therefore, three such exchangers would be reasonable. With carbon steel, 690 kPa design pressure, and an adjustment for tube diameter of ~1.75/0.91 (from Fig. 14-21) and 0.96 for tube length, the base cost is $70,000 for each exchanger.

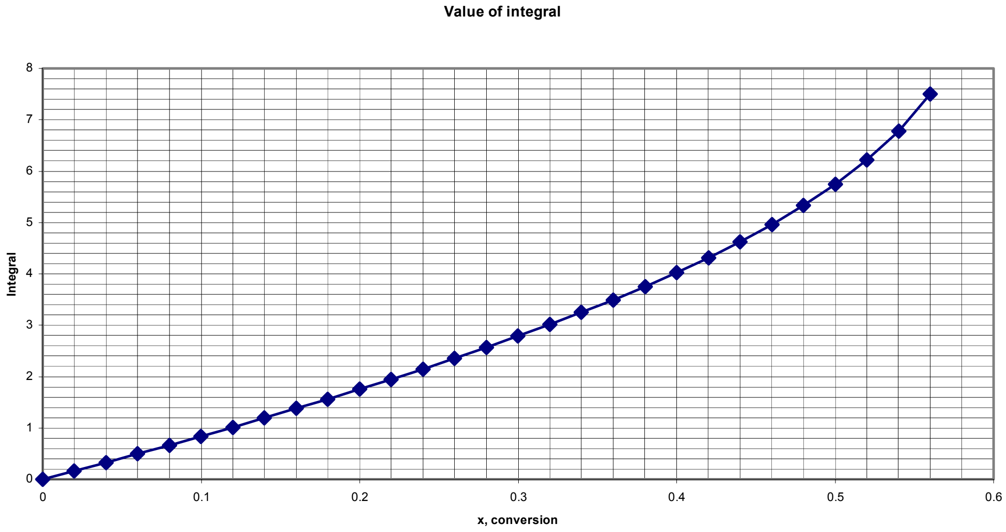
The total cost of the exchangers is estimated to be

(3)($70,000)(1.75/0.91)(0.96) = <u>$388,000</u> <u>ANSWER</u>

It would be expected that the same amount of catalyst as calculated in part b) would still be used. It would be mixed with sufficient inert catalyst support to pack the tubes full.

PROBLEM 13-10 (continued-2)


Rate Integration


Value of integral

PROBLEM 13-11

The time for a batch reaction is given by Eq. 13-8. Since N_i/V_R is the reactant concentration, such as lactose in this case, the equation can be written as

$$\theta = \int_{c_{\text{lactose},i}}^{c_{\text{lactose},f}} \mathrm{d}(c_{\text{lactose}})/(r_{\text{lactose}}) \approx \sum_{j=1}^{j=g} (\Delta c_{\text{lactose}})_j/(r_{\text{lactose}})_{j,\text{average}}$$

where $c_{\text{lactose},f}$ is the final lactose concentration = 20 kg/m^3, $c_{\text{lactose},i}$ the initial lactose concentration = 50 kg/m^3, and $g = (c_{\text{lactose},i} - c_{\text{lactose},f})/\Delta c_{\text{lactose}}$; $\Delta c_{\text{lactose}}$ is negative. Because the rate is a complicated form in concentration, a numerical integration is recommended. Any acceptable numerical integration routine may be used. Here, integration was performed with a spreadsheet using the trapezoidal approximation. The Δ-increment size was decreased in successive integrations until the result does not change significantly with increment size. The term $(r_{\text{lactose}})_{j,\text{average}} = (r_{\text{lactose},j\text{-}1} + r_{\text{lactose},j})/2$ for integration by the trapezoidal method.

The reaction rate is given by

$$r_{\text{lactose}} = -(0.901)\Big(1+(c_{\text{propionic acid}}/4.4214)\Big)^{-1}\Big((c_{\text{lactose}})(c_{P.acidipropionici})/(c_{\text{lactose}} + 32.5)\Big)$$

The concentrations are related as follows:

$$c_{\text{lactose},j} = c_{\text{lactose,i}} + \sum_{j=1}^{j} \Delta c_{\text{lactose}}; \quad c_{\text{propionic acid},j} = -(0.307)\sum_{j=1}^{j} \Delta c_{\text{lactose}};$$

$$c_{P.acidipropionici,j} = 0.1 - (0.145)\,)\sum_{j=1}^{j} \Delta c_{\text{lactose}}$$

The rate as a function of c_{lactose} and the value of the integral as a function of c_{lactose} are shown on the next page.

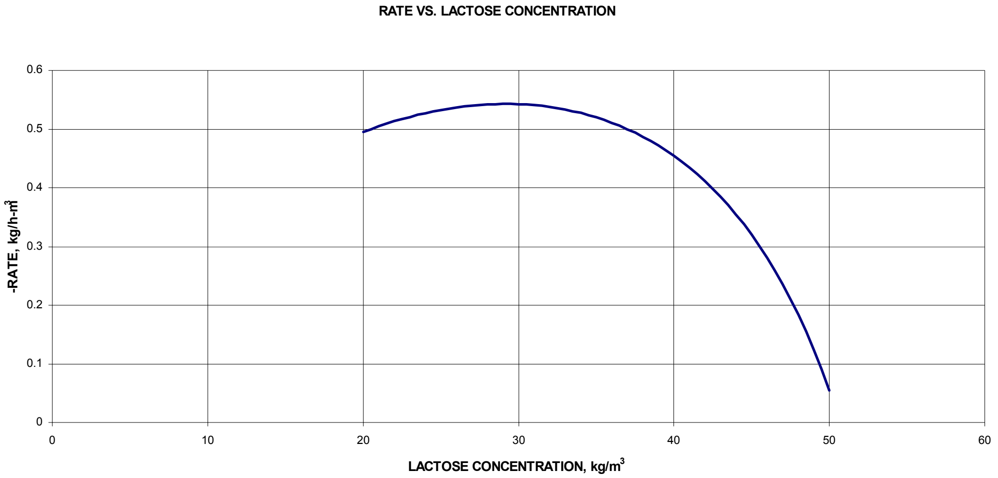
The result for these conditions is

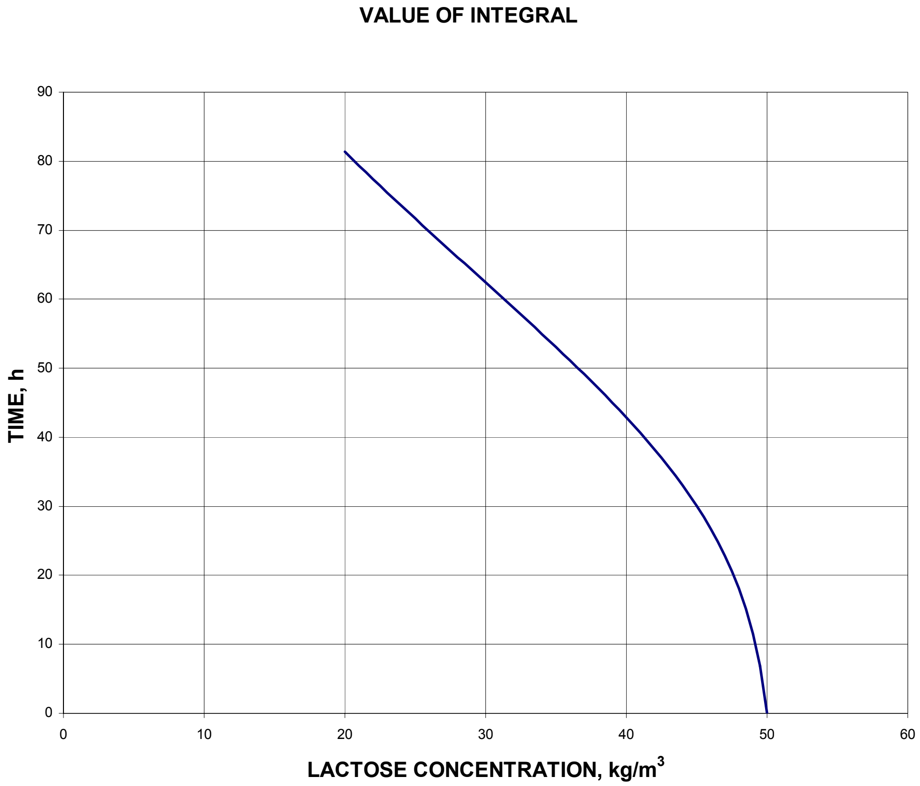
$\underline{\underline{\theta = 81.4 \text{ h}}}$ $\underline{\underline{\text{ANSWER}}}$

The concentrations of lactose, *P. acidipropionici* and propionic acid are shown vs. time in the graph on the second next page.

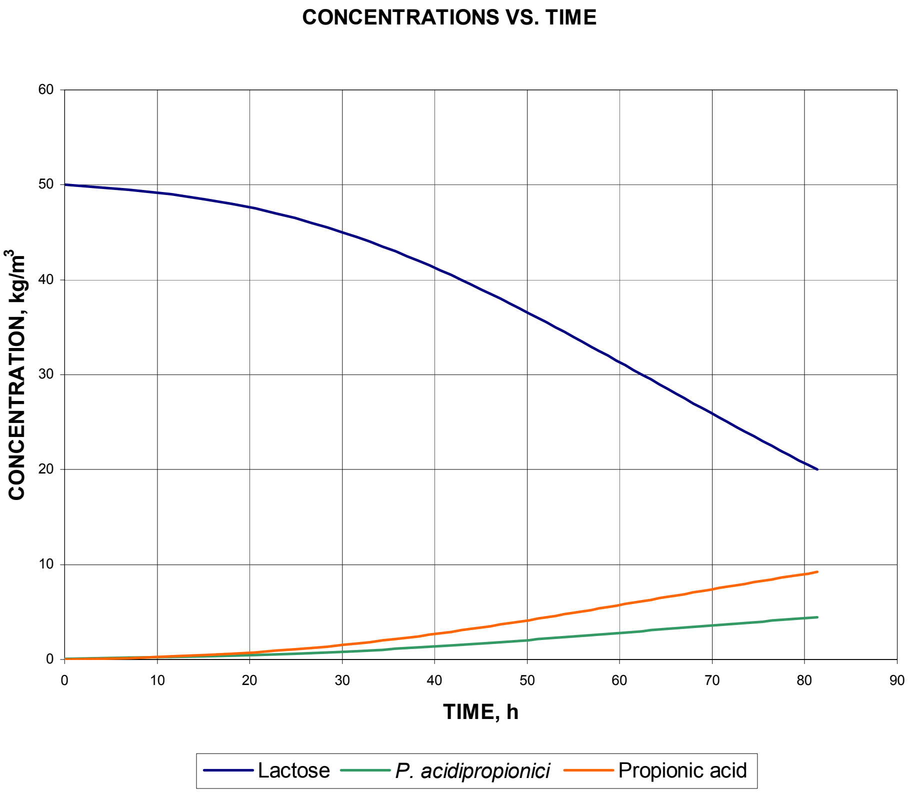
The required reaction time can be substantially reduced by increasing the initial concentration of biomass, *P. acidipropionici*. For example, if the initial concentration is 1 kg/m^3, rather than 0.1 kg/m^3, the time required is 44.4 h.

PROBLEM 13-11 (continued-1)





PROBLEM 13-11 (continued-2)



PROBLEM 13-12

Clarification: The third sentence should read, "In the proposed recovery process, each kg of catalyst-containing ceramic is mixed with two kg of aqueous sodium cyanide solution, and the Pt is dissolved by complexing with the cyanide."

a) Assuming complete Pt recovery and a 90% operating factor, the annual production rate is

$$= (12 \text{ blocks/h})(0.05 \text{ kg-Pt/block})(24*365*0.9 \text{ h/y}) = 4{,}730 \text{ kg-Pt/y}$$

This small rate indicates use of a batch reactor. See Ch. 4.
ANSWER

b) There is no effect of pressure on the reaction, it is liquid phase. So, ambient pressure, or a pressure somewhat lower to prevent cyanide leakage, is recommended.
The temperature is selected to achieve the specified reactor space time of 4 h. Using Eq. 13-10, with θ in s.

$$\theta = (N_{io}/V_R)\int_0^{Xie}[\mathrm{d}X_i/(-r_i)] = c_{io}\int_0^{Xie}\{\mathrm{d}X_i/[(c_{io})(\mathrm{k}_2)(1\text{-}X_i)]\} = (1/\mathrm{k}_2)\ln[1/(1\text{-}X_{ie})]$$

$$= 4*3600 = \{\ln[1/(1\text{-}0.9999)]\}/[(5*10^7)\exp(-13720/\mathrm{T})]$$

from which,

$$\exp(-13720/\mathrm{T}) = \ln(1/0.0001)/[(5*10^7)(4*3600)] = 1.28*10^{-11}$$

$$-13720/\mathrm{T} = -25.08, \text{ and } \mathrm{T} = 547 \text{ K} \quad \text{ANSWER}$$

c) The reactor volume, for a space time of 4 h, is given by Eq. 13-7 as

$$V_R = (F_i*v_{iF})\theta$$

where (F_i*v_{iF}) is equal to the average volumetric flowrate of the reactor feed. This is the flowrate of catalyst blocks plus cyanide solution. Assuming the volume of this mixture is 0.003 m^3/kg of ceramic catalyst (approximated as 1 kg of ceramic per 2 kg of solution, thus ~3 kg/kg of ceramic), the reactor volume is

$$V_R = (12 \text{ blocks/h})(5 \text{ kg ceramic/block})(0.003 \text{ m}^3\text{/kg ceramic})(4\text{h})$$

$$= 0.72 \text{ m}^3 \text{ ANSWER}$$

PROBLEM 13-12 (continued-1)

d) The reactor must contain a hot solution of sodium cyanide/hydrocyanic acid, as well as the abrasive ceramic material of the catalyst block. Either a steel or glass-lined reactor might be used. Steel has the problem of being somewhat subject to corrosion by weak acids, while a glass-lined vessel has the problem of being subject to scratching and erosion from the ceramic catalyst support. Costs are estimated from Fig. 13-15 to be

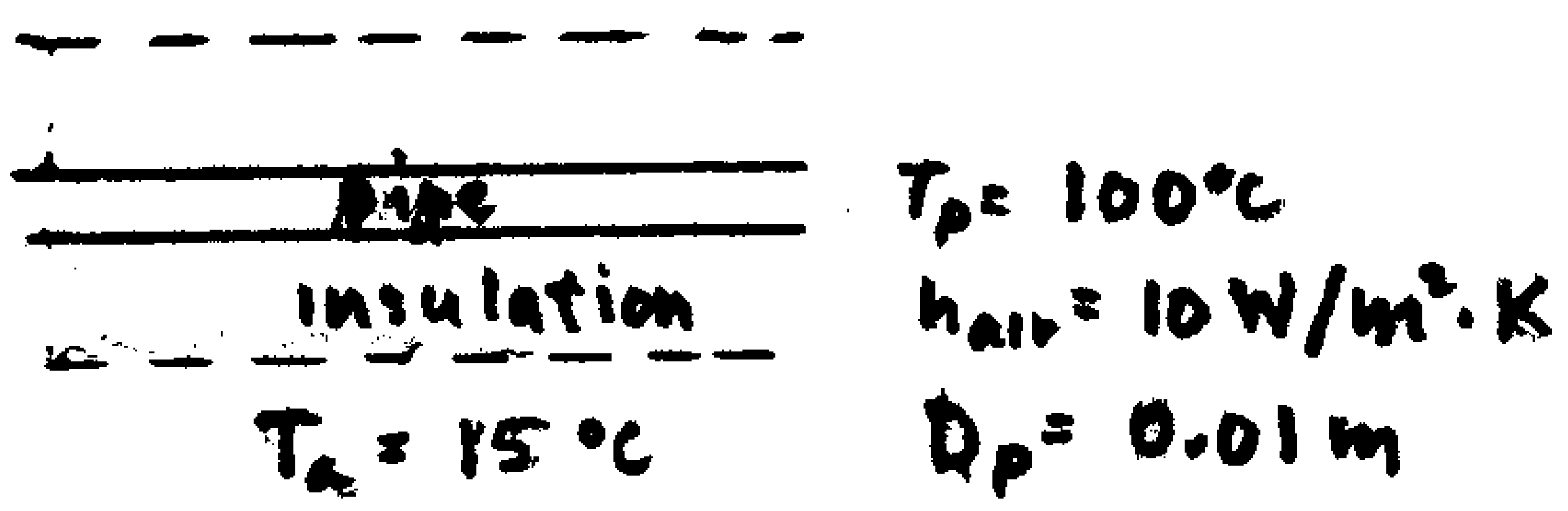
Glass-lined-reactor cost = $24,000
Steel reactor cost = $10,000 (TENTATIVE ANSWERS)

Further study, including more detailed corrosion and erosion information and an alternative investment comparison, would be needed to make a definitive recommendation.

14-1

T_p = temperature on outside of pipe

T_a = temperature of ambient air



Assume that the thermal resistance on the fluid side of the pipe can be neglected. First consider the heat transfer per meter of uninsulated pipe. For the latter, the thermal resistance is

$$R_{T_a} = \frac{T_p - T_a}{\dot{q}} = \frac{1}{2\pi r_p hL} = \frac{1}{2\pi(0.005)(10)(1)} = 3.183(\mathrm{K/W})/\mathrm{m}$$

The rate of heat loss per meter of pipe is then

$$\dot{q} = \frac{T_p - T_a}{R_{T_a}} = \frac{100-15}{3.183} = \underline{\mathbf{26.70 W/m}}$$

Answer

The thermal resistance of the first layer of insulation is given by

$$R_{T_1} = \frac{\ln r_1 / r_p}{2\pi L k} = \frac{\ln(0.01/0.005)}{2\pi(1)(0.1)} = 1.1032(\mathrm{K/W})/\mathrm{m}$$

where r_1 = radius of pipe plus one layer of insulation.

Because of a doubling of the surface area of the insulation, the thermal resistance becomes half of that for the uninsulated pipe as given by

$$R_{T_a} = \frac{1}{2\pi r_1 hL} = \frac{1}{2\pi(0.01)(10)(1)} = 1.592(\mathrm{K/W})/\mathrm{m}$$

The total thermal resistance is

$$R_T = 1.1032 + 1.592 = 2.695(\mathrm{K/W})/\mathrm{m}$$

and the heat flow is

$$\dot{q} = \frac{100-15}{2.695} = 31.54\ \mathrm{W/m}$$

Note that adding 0.005m of insulation increased the heat loss.

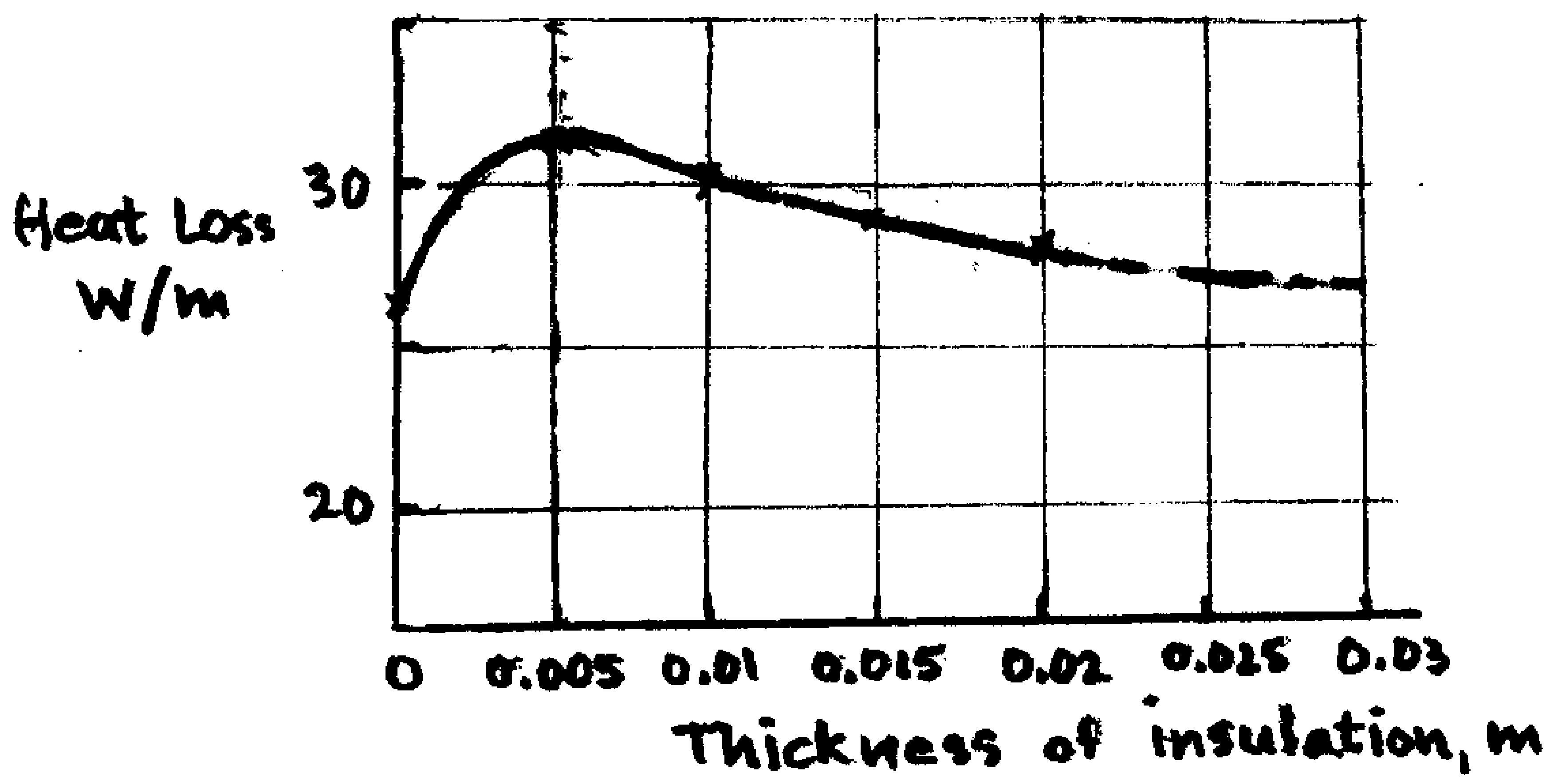
14-1 (continued)

The addition of another 0.005m of insulation on the pipe results in

$$R_{T_2} = 1.748\text{(K/W)/m} \quad R_{T_a} = 1.061\text{(K/W)/m}, \quad R_T = 2.809\text{(K/W)/m}$$

and the heat loss is 30.25W/m

These calculations are repeated with the addition of each layer of insulation. The results are shown below.



For small diameter pipes there is a critical thickness of insulation that produces the minimum thermal resistance. To be effective insulation thicknesses must be greater than this value. In this problem the critical thickness of insulation is 0.005m as shown on the figure. This can be verified by differentiating the sum of the conductive and convective resistance with respect to the outside radius and setting the result equal to zero. Thus

$$R_T = \frac{\ln r_2 / r_p}{2\pi Lh} + \frac{1}{2\pi r^2 Lk}$$

$$\frac{dR_T}{dr_2} = 0 \quad \text{and} \quad r_2 = k/h = 0.1/10 = 0.1\text{ m}$$

In practice, the optimum thickness of insulation involves an economic balance between additional costs for the insulation and the energy savings attributed to the insulation. In this problem, it requires adding 0.02m of insulation to reduce the heat losses per meter of pipe to the heat losses shown for the uninsulated pipe. Since additional insulation only reduces the losses by small amounts, it is more economical to operate the pipe without the addition of any insulation. This is often the case for small pipes where the heat loss is relatively small. **Answer**

14-2

$$T = 350\text{K} = 77^\circ\text{C} \qquad \text{Re} = 5\times10^4 \qquad \text{ID} = 0.0254\text{m}$$

For the same Reynolds number, the average velocity will be different for each fluid

$$V = \frac{\text{Re}\,\mu}{D\rho} \quad \text{for air} \quad V = \frac{(5\times10^4)(2\times10^{-5})}{(0.0254)(0.955)} = 41.22\,\text{m/s}$$

$$\text{for H}_2\text{O} \quad V = \frac{(5\times10^4)(3.72\times10^{-4})}{(0.0254)(973)} = 0.75\,\text{m/s}$$

$$\text{for oil} \quad V = \frac{(5\times10^4)(3.56\times10^{-2})}{(0.0254)(854)} = 83\,\text{m/s}$$

Use Eq. (14-18) to obtain the individual heat transfer coefficient. Assume that $(\mu/\mu_w)^{0.14}$ is essentially unity.

$$h = (k/D)(0.023)\,\text{Re}^{0.8}\,\text{Pr}^{1/3}$$

$$h_{\text{air}} = (0.030/0.254)(0.023)(5\times10^4)^{0.8}[(1050)(2\times10^{-5})(0.030)]^{1/3} = 136\ \text{W/m}^2\cdot\text{K}$$

$$h_{H_2O} = (0.668/0.0254)(0.023)(5\times10^4)^{0.8}[(4190)(3.72\times10^{-4})(0.668)]^{1/3} = 4610\ \text{W/m}^2\cdot\text{K}$$

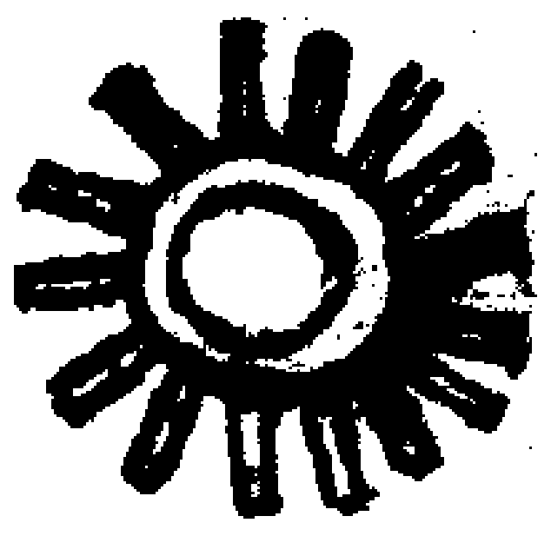
$$h_{\text{oil}} = (0.138/0.0254)(0.023)(5\times10^4)^{0.8}[(2116)(3.56\times10^{-2})(0.138)]^{1/3} = 5860\ \text{W/m}^2\cdot\text{K}$$

The term $\rho V^2/2$ for each fluid, because of the constant Reynolds number, provides an indication of the magnitude of the pressure loss in the pipe. The values for air, water, and oil are 811 N/m^2, 273 N/m^2, and 2.94×10^6 N/m^2, respectively.

Summary	Air	Water	Oil
Velocity, m/s	41.2	0.75	83
Heat transfer coeff, W/m$^2\cdot$K	136	4610	5860
Relative pressure drop, N/m^2	811	273	2.94×10^6

Obviously the pressure drop for the oil is too large and the velocity of the oil must be reduced to a reasonable value. However, in reducing the velocity, the flow may become laminar with a large decrease in the heat transfer coefficient. This indicates that heat transfer rates for oil that are equivalent to those for water cannot be achieved with pressure drops similar to those obtained with water flow.

14-3



$D_2 = 0.022\text{m} \quad D_1 = 0.028\text{m} \quad L_p = 0.012\text{m} \quad t_s = 0.002\text{m}$

Inside surface areas/unit length of pipe, $A_2 = \pi D_2 = \pi(0.022) = 0.069\,\text{m}^2/\text{m}$

Outside surface area (without fins)/unit length of pipe, $A_1 = \pi(0.028) = 0.0879\,\text{m}^2/\text{m}$

Outside surface area not covered with fins/unit length of pipe,

$$A_p = 0.0879 - \pi(0.028)(0.002)/0.01 = 0.0703\,\text{m}^2/\text{m}$$

Surface area of fin/unit length of pipe, $A_f = \dfrac{2\pi[(0.04)^2 - (0.028)^2}{4(0.01)} + \dfrac{\pi[(0.04)(0.002)}{0.01}$

$$= 0.1533\,\text{m}^2/\text{m}$$

Total outside surface area, $A_T = 0.1533 + 0.0703 = 0.2236\,\text{m}^2/\text{m}$

We can use Eq. (14-26) to obtain the fin efficiency. For a quick approximation, use the approach suggested in Chap. 2.5.3 of the *Heat Exchanger Design Handbook.*

Effective fin height $L_e = (L_f + t_f)/2 = (0.012 + 0.002)/2 = 0.007\,\text{m}$

Fin parameter $M = (h/k_f t_f/2)^{1/2} = [200/(25)(0.001)^{1/2} = 89.44$

$$\eta_f = [1 + 1/3(ML_e)^2(D_f/D_o)^{1/2}]^{-1}$$
$$= \{1 + 1/3[(89.44)(0.007)]^2[(0.040/0.028)]^{1/2}\}^{-1} = 0.865$$

Since no fouling factors were given, assume the heat transfer occurs when the heat exchange process has just initiated operation. Calculate the effective outside heat transfer coefficient.

$$h_e = (h_o/A_o)(A_f\eta_f + A_p)$$
$$= (200/0.2236)[(0.1533)(0.865) + 0.0703] = 181.5\ \text{W/m}^2\cdot\text{K}$$

This permits calculation of the overall heat transfer coefficient based on total outside area (fins + outside area not covered by fins)

14-3 (continued)

$$\frac{1}{U_o} = \frac{A_o}{h_i A_1} + \frac{D_1 \ln D_1 / D_2}{2k_w} \frac{A_o}{A_1} + \frac{1}{h_e'}$$

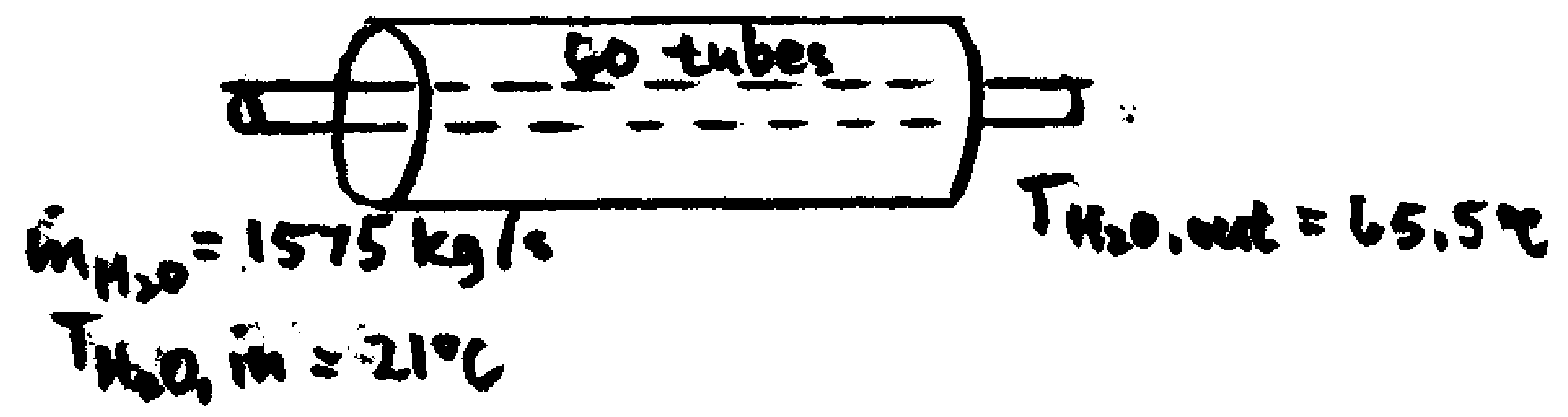
$$= \frac{0.2236}{(1500)(0.069)} + \frac{(0.028)\ln(0.028/0.022)}{(2)(25)} \frac{(0.2236)}{0.0879} + \frac{1}{181.5}$$

$$= 2.160\times10^{-3} + 3.434\times10^{-4} + 5.5096\times10^{-3} = 8.013\times10^{-3}$$

$$U_o = \underline{\underline{125\ \text{W/m}^2\cdot\text{K}}}$$

If the heat transfer coefficient is based on the outside area of the pipe, A_1, as it is used in Eq. (14-29), the value for the coefficient would be $\underline{\underline{318\ \text{W/m}^2\cdot\text{K}}}$. Thus, great care must be exercised to define what area is used as a reference point when fins are involved.

14-4 Tube and shell heat exchanger



h_{H_2O} + with scale $= 8250\ \text{W/m}^2\cdot\text{K}$

$h_{\text{steam}} = 11{,}400\ \text{W/m}^2\cdot\text{K}$

$T_{1\ \text{steam}} = 143.3°\text{C}$

$\rho_{av,H_2O} = 988.5\ \text{kg/m}^3$

$K_{\text{steel}} = 45\ \text{W/m}^2\cdot\text{K}$

Case 1 = Design with scale, $h_{o,H_2O} = 8250\ \text{W/m}^2\cdot\text{K}$

Case 2 = Design with no scale

$\dot{q} = \dot{q}_2 \quad \dot{q}_1 = U_1 A_1 \Delta T_1 \quad \dot{q} = U_2 A_2 \Delta T_2 \quad A_1 = A_2 \quad \Delta T_2 = (U_1 / U_2)\Delta T_1$

Base both U_1 and U_2 on the inside area

$$\Delta T_{1,\,\text{avg}} = \frac{(143.3-21)+(143.3-65.5)}{2} = 100°C \qquad \Delta T_{2,\,\text{avg}} = \frac{(T_{2,s}-21)+(T_{2,s}-65.5)}{2} = T_{2,s} - 43.3$$

Determine U based on average values for physical properties

$$V_{1,\text{avg}} \left(\frac{15.75}{60}\frac{\text{kg}}{\text{s}}\right)\left(\frac{\text{m}^3}{988.5\,\text{kg}}\right)\left(\frac{4}{\pi(0.0186)^2\text{m}^2}\right) = 0.977\ \text{m}^3/\text{s}$$

$$\text{Re}_1 = DV\rho/\mu_{\text{avg}} = (0.0186)(0.977)(988.5)/0.000616 = 29{,}160 \text{ turbulent flow}$$

$$h_i = (k/D_i)(0.023)\text{Re}^{0.8}\,\text{Pr}^{1/3} \text{ neglecting viscosity correction}$$

$$=(0.636/0.0186)(0.023)(29{,}160)^{0.8}\left[\frac{(4.179)(0.000616)}{0.636}\right]^{1/3} = 4678\ \text{W/m}^2\cdot\text{K}$$

From Eq. (14-4a)

$$\frac{1}{U_1} = \frac{1}{4678} + \frac{1}{8250} + \frac{\pi(0.0186)(0.0034)}{(45)(0.0686)} + \frac{(0.0186)}{(11400)(0.0254)} = 4.636\times10^{-4}\ \text{m}^2\cdot\text{K/W}$$

$$U_1 = 2157\ \text{W/m}^2\cdot\text{K} \qquad \text{where } A_{m,W} = 0.068\ \text{m}^2$$

$$\frac{1}{U^2} = 4.636\times10^{-4} - \frac{1}{h_D} = 4.636\times10^{-4} - \frac{1}{8250} = 3.424\times10^{-4}\ \text{m}^2\cdot\text{K/W}$$

$$U_{2_1} = 2920\ \text{W/m}^2\cdot\text{K}$$

14-4 (continued)

From above

$$\Delta T_2(U_1/U_2)\Delta T_1 = (2157/2980)(100) = 72.4^\circ\text{C}$$

$$\Delta T_{2,\text{avg}} = 72.4 = T_{2,\text{steam}} - 43.3 \quad \text{if } \Delta T_1/\Delta T_2 < 2.0 \text{ in case 2}$$

$$T_{2,s} \quad = 72.4 + 43.3 = 115.7^\circ\text{C} \qquad \Delta T_1 = 115.7 - 21 = 94.7^\circ\text{C}$$

$$\Delta T_2 = 115.7 - 65.5 = 50.2^\circ\text{C}$$

$$\Delta T_1/\Delta T_2 = 94.7/50.2 < 2.0$$

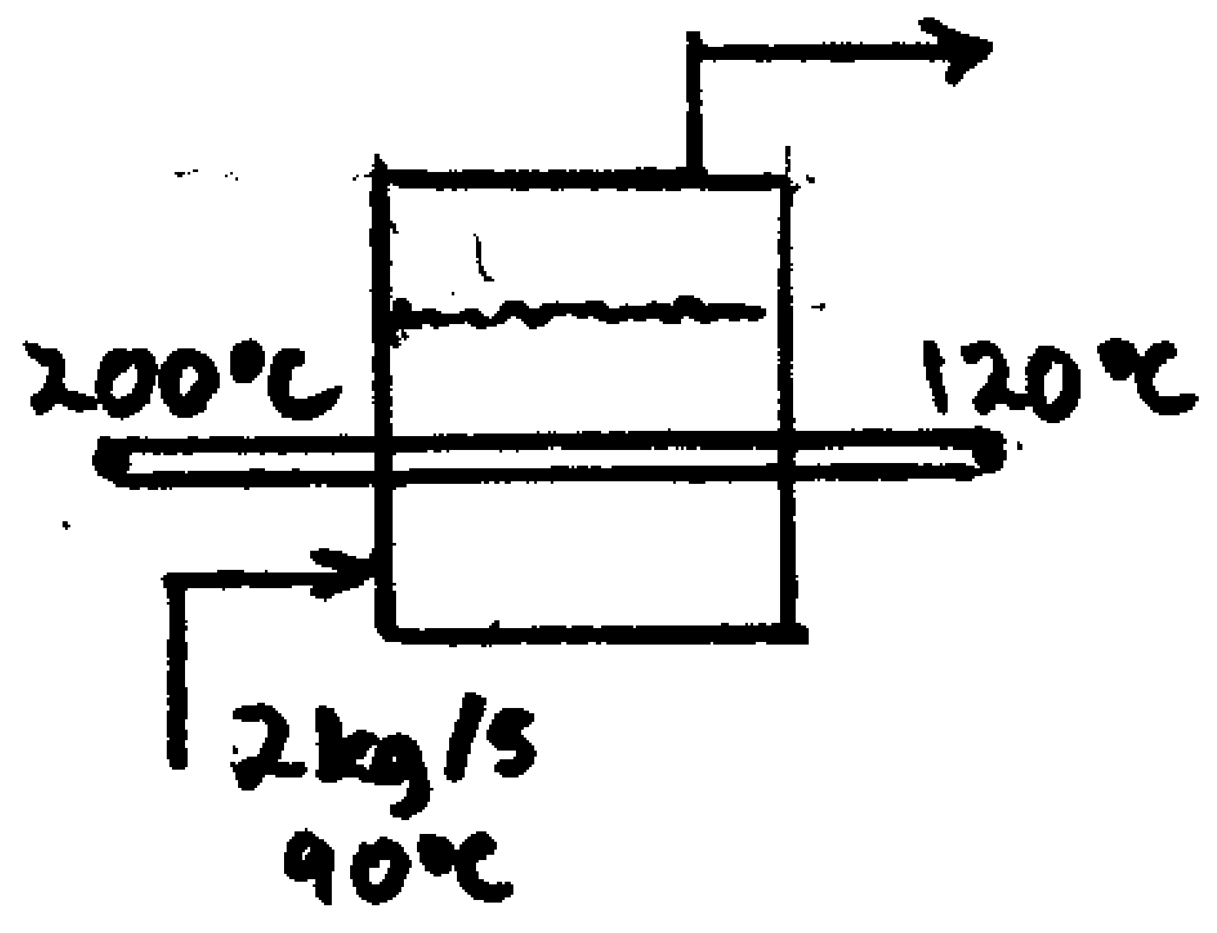
thus, using averages is acceptable

∴ Temperature of steam required is **115.7°C or 116°C**

Answer

14'.8

14-5



$\Delta H_v = 200\,\text{kJ/kg} @ 90^\circ\text{C}$

$\dot{m}_c = 2\,\text{kg/s}$

$T_{h,1} = 200^\circ\text{C} \quad C_{p,h} = 2.2\,\text{kW/kg}\cdot\text{K}$

$T_{h,2} = 120^\circ\text{C}$

$U_o = 400\,\text{W/m}^2\cdot\text{K} \quad \text{Find}\, \dot{m}_h, \Delta T_{\ln m}, A$

The evaporization process is represented by Fig. 14-2c. Since C_p is constant for the hot fluid, a $\Delta T_{\ln m}$ can be determined from

$$\Delta T_m = \frac{(200-90)-(120-90)}{\ln 110/30} = \frac{80}{1.299} = 61.6^\circ\text{C}$$

$$\dot{q}_c = \dot{m}_c \Delta H_V = (2)(200{,}000) = 400{,}000\,\text{J/s} \quad \dot{q}_c = \dot{q}_h$$

$$\dot{q}_h = \dot{m}_h Cp_h (T_{h,m} - T_{h,out})$$

$$\dot{m}_h = \dot{q}_h / Cp_h (T_{h,m} - T_{h,out}) = -400{,}000/(2{,}200)(120-200)$$

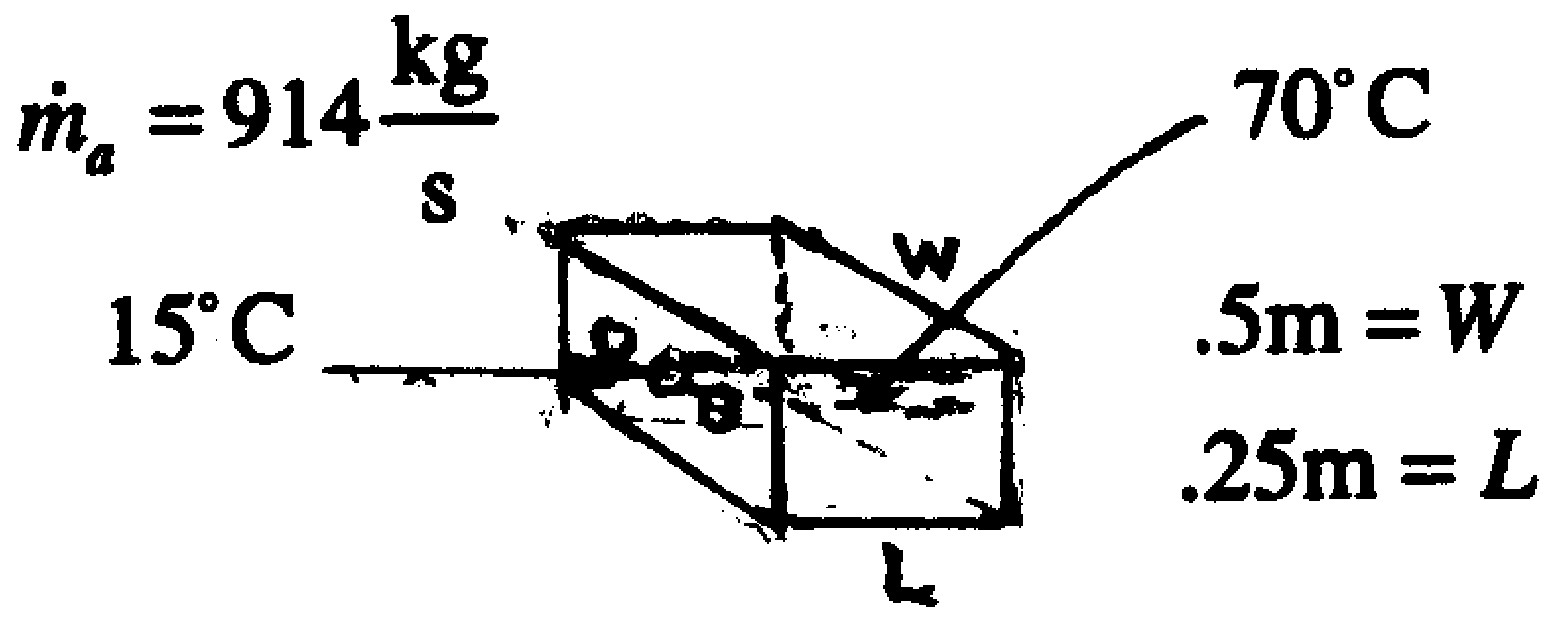
$$= \underline{\underline{2.27\ \text{kg/s}}}$$

Answer

$$A = \dot{q} / U \Delta T_{em} = 400{,}000/(400)(61.6) = \underline{\underline{16.2\text{m}^2}}$$

Answer

14-6



$D_o = 0.0104\text{m}$
$L_p = 1.0\text{m}$

staggered array with 60° triangular pitch
56 tubes 7x8 with a pitch of 0.313m

Fluid properties of air at 15°C

$\mu = 1.79\times10^{-5}\,\text{Pa}\cdot\text{s} \quad k = 2.53\times10^{-2}\,\text{W/m}\cdot\text{k} \quad Cp = 1007\,\text{J/kgK}$

$\rho = 1.226\,\text{kg/m}^3$

Calculate the maximum air velocity from

$$V_{\max} = \frac{m}{\rho n(P-D)L}$$

where n = number of tubes in a row
P = pitch of 0.0313 m

$$= \frac{0.914}{(1.226)(8)(0.0313-0.0104)(0.5)} = 8.92\,\text{m/s}$$

The corresponding Reynolds number is

$$\text{Re} = DV_{\max}\rho/\mu = (0.0104)(8.92)(1.226)/1.79\times10^{-5} = 6350$$

Use Eq. (14-22) to obtain average heat transfer coefficient

$$\frac{\bar{h}_o D_o}{K} = a\,\text{Re}^m\,\text{Pr}^{0.34}\,F_1F_2$$

Table 14-1 provides a = 0.273, m = 0.635. Since the difference between the bulk air temperature and the pipe surface temp is small, $F_1 \cong 1$. Table 14-2 indicates that $F_2 = 0.97$ for a staggered array of seven rows. Thus,

14-6 (continued)

$$\bar{h}_o = (2.53\times10^{-2}/0.0104)(0.273)(6350)^{0.635}\left[\frac{(1007)(1.79\times10^{-5})}{2.53\times10^{-2}}\right]^{0.34}(0.97)$$

$$= 149.2\,\mathrm{W/m^2\cdot K}$$

Now calculate the log mean temperature difference in terms of the average heat transfer coefficient

$$\Delta T = \frac{(T_p - T_1) - (T_P - T_2)}{\ln\left[(T_p - T_1) - (T_P - T_2)\right]}$$ this can also be written in terms of $\bar{h}_o$ as

$$= (T_p - T_1)\left[\frac{1 - e^{-A\bar{h}_o/\dot{m}C_p}}{A\bar{h}_o/\dot{m}C_p}\right]$$ $A = \pi DLN_T$ where N_T = number of tubes

$$A = \pi(0.0104(1)(56) = 1.829\,\mathrm{m}^2$$

$$A\bar{h}_o/\dot{m}C_p = (1.829)(149.2)/(0.914)(1007) = 0.296$$

$$\Delta T = (70 - 15)\left(\frac{1 - e^{-0.296}}{0.296}\right) = 47.6^\circ\mathrm{C}$$

The total heat transfer rate is then

$$\dot{q} = \bar{h}A\Delta T = (149.2)(1.829)(47.6) = 12{,}989\,\mathrm{W} \cong 13{,}000\,\mathrm{W}$$

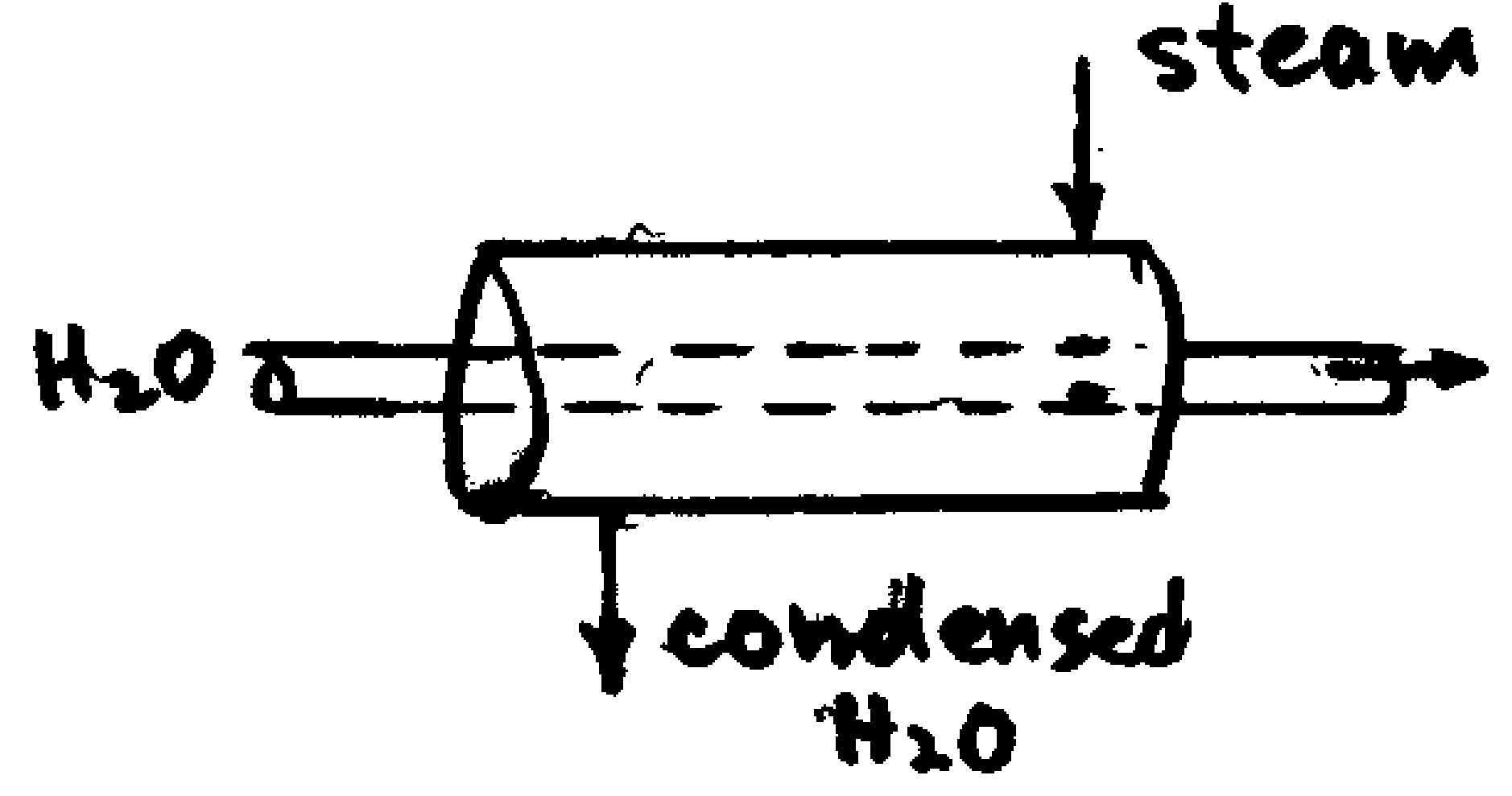
This corresponds to an increase in the air temperature of

$$T_{exit} - T_{inlet} = \dot{q}/\dot{m}C_p = 12{,}989/(0.914)(1007) = 14.1^\circ\mathrm{C}$$

and an exit temperature of 14.1 + 15 = **29.1°C**

Answer

14-7



turbulent flow

$\dot{m}_{H_2O}, T_{H_2O,in}, T_{H_2O,out}, T_s, \Delta p_{tube}$ assumed to be constant

assume negligible thermal resistance of tube and for steam condensate film, tube diameter is available

$\dot{q} = U_i A_i \Delta T_{lm} = \text{constant}/U_i$, From Eq. (14-14a),

$$\frac{1}{U_i} = \frac{1}{h_i} + \frac{A_i x_w}{k_w A_{w,m}} + \frac{D_i}{h_o D_o}$$

assume last two terms are negligible

From Eq. (14-18)

$$h_i = 0.023\ \text{Re}_i^{0.8}\,\text{Pr}_i^{1/3}(\mu/\mu_w)^{0.14}$$

assume $(\mu/\mu_w)^{0.14} \sim 1.0$

$$h_i \cong \text{Re}_i^{0.8} \cong K(DV\rho/\mu)^{0.8} \qquad V = \dot{m}/(\rho\pi D_i^2/4) \qquad V \cong 1/D_i^2/4$$

$$\cong (D_i/D_i^2)^{0.8} \cong 1/D^{0.8} \cong D^{-0.8}$$

$$\frac{1}{U_1} \cong \frac{1}{D_i^{-0.8}}$$

$$A_i \cong D_1^{0.8}$$

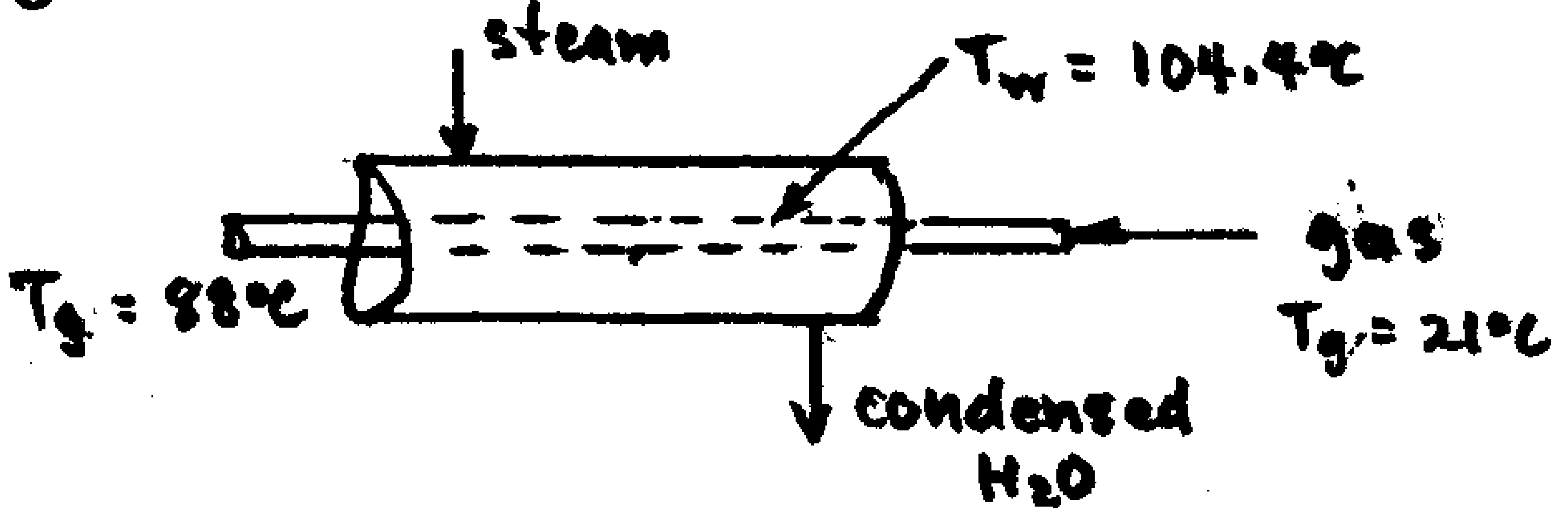
Answer

Assuming flow is turbulent

Relative D_i	Relative A_i
1	1
2	1.74
3	2.41
4	3.02
5	3.62
10	6.31

14', 12

14-8



Assume that the gas has the properties of air. These properties, particularly density, depend on the pressure of the gas. Since such information was not included, consider two cases: one with a pressure of 101 kPa and the other with a pressure of 202 kPa for the gas. At 101 kPa, the properties for the air are:

$$\rho = 1.078\,\mathrm{kg/m^3} \quad C_p = 1046\,\mathrm{J/kg\cdot K} \quad k = 0.0282\,\mathrm{W/m\cdot K} \quad \mu = 0.018\,\mathrm{cP}$$

Determine the optimal velocity from a rearrangement of Eq. (13-15) assuming turbulent flow and $D_i = 0.023$.

For a pressure of 101 kPa

$$V^{0.45} = \left[D_i/(0.363)\right] A^{0.45}\rho^{0.13}$$

$$V^{0.45} = (0.023)/(0.363)[(0.785)(0.023)^2]^{0.45}(1.078)^{0.13} = 2.09$$

$$V \cong 5.2\,\mathrm{m/s}$$

$$V = \dot{m}/\rho A_c$$

Since $\dot{m}_{tube} = V\rho A_c = (5.2)(1.078)0.785)(0.023)^2 = 2.3278\times10^{-3}\,\mathrm{(kg/s)/tube}$

$$N_T = \dot{m}/\dot{m}_{tube} = 0.075/2.3278\times10^{-3} = \underline{\underline{32.2 \text{ tubes} \sim 32 \text{ tubes}}}$$

Answer

The heat transferred to the gas is

$$\dot{q}_i = \dot{m}\,C_p\,\Delta T_{gas} \qquad \Delta T_{\ln m} = \frac{(104.4-21)-(104.4-88)}{\ln 83.4/16.4}$$

$$= (0.075)(1046)(88-21) \qquad = 41.2^\circ\mathrm{C}$$

$$= 5260\,\mathrm{J/s}$$

$$\dot{q}_i = A_i U_i\,\Delta T_{\ln m}$$

From Eq. (14-4a)

14-8 (continued)

$$\frac{1}{U_1}=\frac{1}{h_i}+\frac{A_i x_W}{k_w A_{w,m}}+\frac{D_i}{h_o D_o}$$

assume that the thermal resistance of the pipe and the steam film are negligible

From Eq. (14-18)

$$h_i=(k/D_i)(0.023)(\mathrm{Re})^{0.8}(\mathrm{Pr})^{1/3}(\mu/\mu_W)^{0.14}$$

$$=\left(\frac{0.0282}{0.023}\right)(0.023)\left[\frac{(0.023)(5.2)(1.078)}{0.000018}\right]^{0.8}\left[\frac{(1046)(0.000018)}{0.0282}\right]^{1/3}$$

$$=29.9 W/\mathrm{m}^2\cdot\mathrm{K}\cong 30\ \mathrm{W/m^2\cdot K}$$

$$L=(\dot{q}/N_T)\pi D_i \mu_i \Delta T_{\ln m}$$

$$=(5260/32)(\pi(0.023)(30)(41.2)$$

$$=\underline{\underline{1.84\,\mathrm{m\ or\ }6.0\,\mathrm{ft}}}$$

Answer

If the gas pressure is 202 kPa, $\rho = 2.1545\,\mathrm{kg/m^3}$ and using the same procedure as above

$V = 4.25\,\mathrm{m/s}$

$N_T = 19.7\ \mathrm{tubes} \cong \underline{\underline{\mathbf{20\ tubes}}}$

Answer

$\mathrm{h_i} = 44.3\,\mathrm{W/m^2\cdot K}$

$\mathrm{L} = \underline{\underline{\mathbf{2.13\ m = 7\ ft}}}$

Answer

This procedure can be followed to obtain appropriate values for other gas conditions.

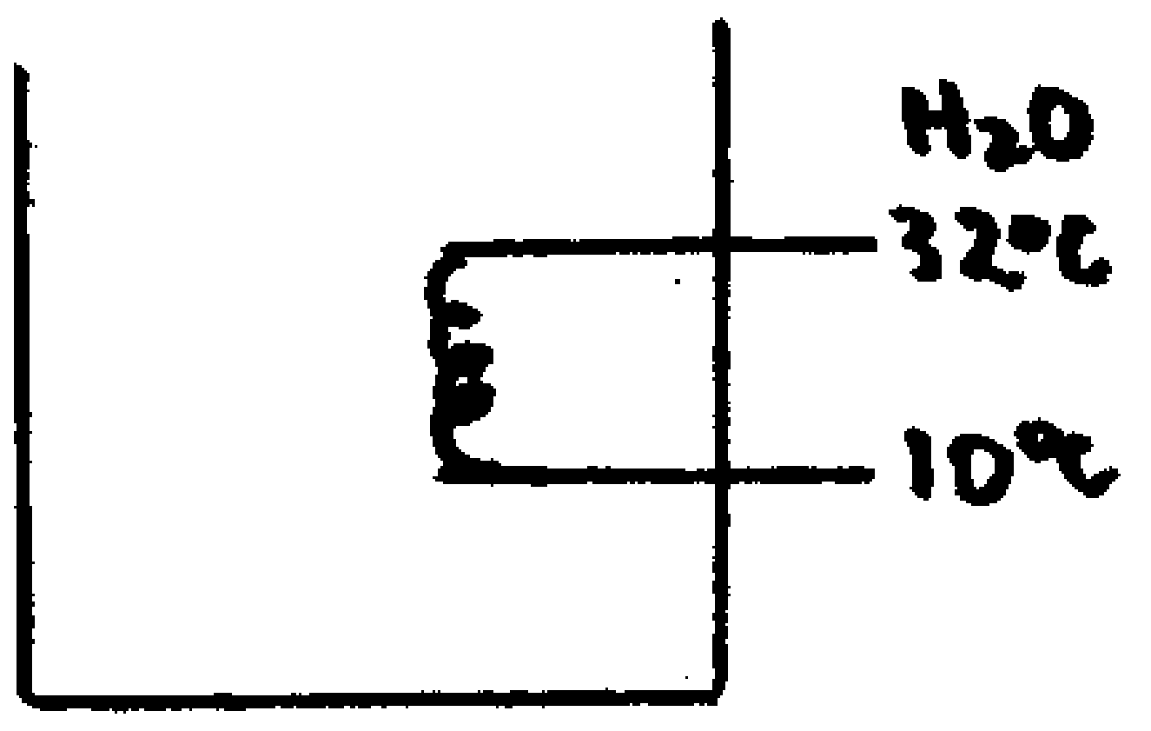
14-8 (continued)

Consider the two cases for providing a 20 percent safety factor.

1. Adding 20% more tubes of the same diameter and length increases the heat transfer area by 20%, but the mass flow in each tube is now 20% less and the velocity in each tube is reduced by 20%. This reduces the heat transfer coefficient inside the tubes by approximately 16% which will require about a 20% increase in heat transfer area to obtain the desired heat transfer. Thus, the desired safety factor will not be achieved by this action.
2. Increasing the tube diameter by 20 percent while keeping the same number fo tubes and same tube length, increases the heat transfer area by 55% and reduces the velocity in each tube by 30.6% and the heat transfer coefficient by 25.3%. The increase in diameter of the tubes provides approximately a 7.5% safety factor. Again the desired safety factor will not be achieved.

A better choice to obtain a 20% safety factor is to increase the length of the tubes by 20%. This will maintain the same mass flow and velocity in each tube. The heat transfer coefficient will not change and 20% more heat can be transferred. The disadvantage of this approach is that the pressure drop across the exchanger will be increased by 20%. If this exceeds the allowable limit, then additional changes in tube size and number will need to be taken.

14-9



$\Delta H_{\text{cond}} = 335\,kJ/\text{kg at }88^\circ\text{C}$ $\quad \dot{m}_{\text{cond}} = 0.126\,\text{kg/s}$

$h_{\text{cond}} = 1420\,W/\text{m}^2\cdot\text{K}$ $\quad k_{\text{pipe}}\ @\ 21^\circ\text{C}{=}385\ \text{W/mK}$

$\text{D}_{\text{i}} = 0.0127\,\text{m}$

$D_o = 0.0152\,m$

Heat requirement is

$$\dot{q} = (0.126)(335) = 42.2\,\text{kJ/s} = \dot{m}_{\text{H}_2\text{O}} C_{p_{\text{H}_2\text{O}}}(32-10)$$

$$\dot{m}_{\text{H}_2\text{O}} = 42.2/(4.186)(22) = 0.458\,\text{kg/s}$$

$$V_{\text{H}_2\text{O}} = (0.485/997)/0.785(0.0127)^3 = 3.63\text{m/s}$$

$$\text{Re}_{H_2O} = \frac{DV\rho}{\mu} = \frac{(0.0127)(3.63)(997)}{0.000983} = 46,750 \quad \text{flow is turbulent}$$

Obtain heat transfer coefficient for the water using Eq. (14-18) with physical properties at the average temperature of the water, namely 21°C.

$$h_i = (0.023)\left(\frac{0.604}{0.0127}\right)(46,750)^{0.8}\left[\frac{(4186)(0.000983)}{0.604}\right]^{1/3} = 11,110\,\text{W/m}^2\cdot\text{K}$$

Base the overall heat transfer coefficient on the inside area using Eq. (14-4a)

$$\frac{1}{U_i} = \frac{1}{11,110} + \frac{(0.0127)\ln(0.0152/0.0127)(0.0025)}{(0.0152-0.0127)(385)} + \frac{0.0127}{(1420)(0.0152)}$$

$$= 6.843\times10^{-4}$$

$$U_i = 1461\ \text{W/m}^2\cdot\text{K}$$

$$\dot{q} = U_i A\Delta T_{\text{avg}} = 42.2\,\text{kJ/s}$$

$$\Delta T_{\text{avg}} = [(88-15)+(88-32)]/2 = 64.5^\circ\text{C}$$

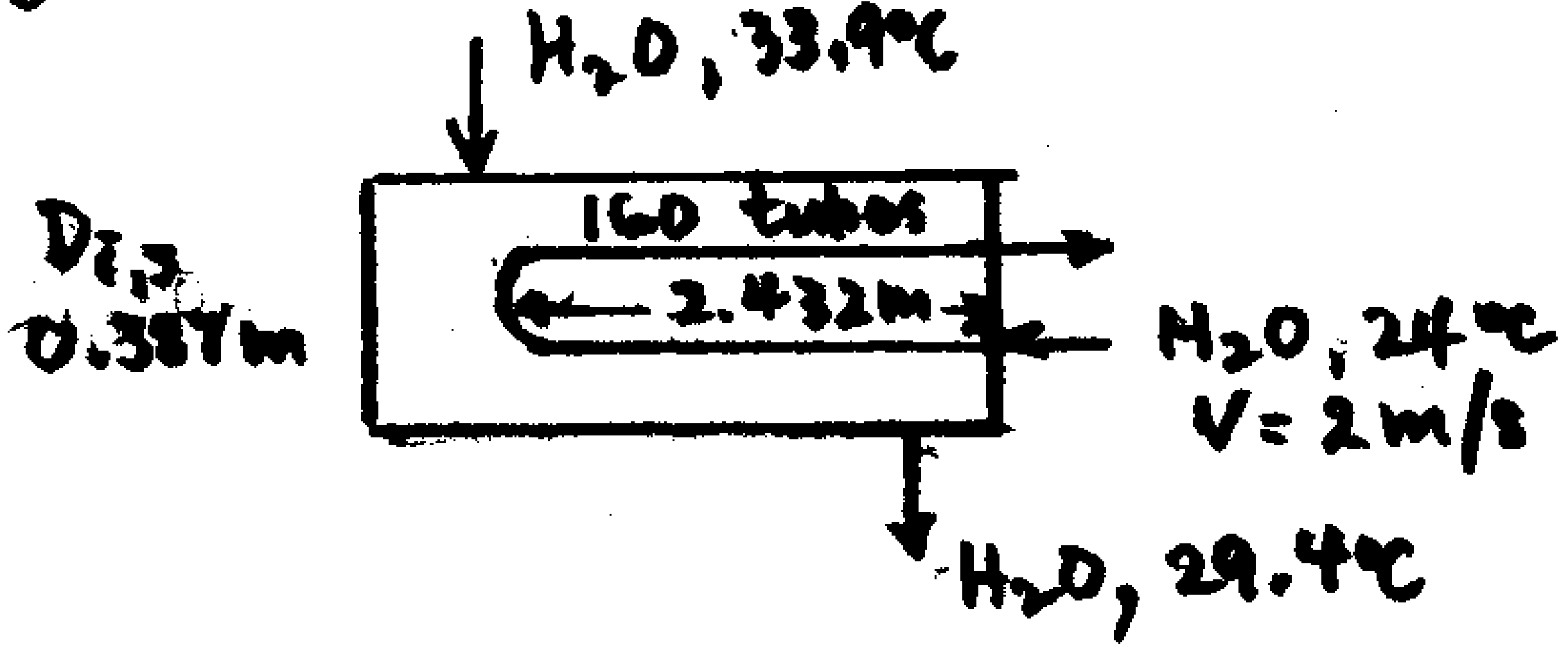
$$\text{A}{=}\pi\text{D}_{\text{i}}L = \dot{q}/U_i\Delta T_{\text{avg}}$$

$$L = \dot{q}/\pi D_i U_i \Delta T_{\text{avg}} = (42,200)(6.843\times10^{-4})/\pi(0.0127)(64.5)$$

$$L = \underline{\underline{11.2\,\text{m}}}$$

Answer

14-10



$D_o = 0.0191$ m, 18BWG $t_W = 0.0025$m

$V = 2.0$m/s $D_i = 0.0166$m

Δp_i and $\Delta p_o < 69$kPa

$h_{o,d}$(distilled water) = 11,360 W/m^2·K $h_{i,d}$(cooling water) = 5680 W/m^2·K

From a heat balance on distilled water side

$$\dot{q} = -(22)(4179)(33.9 - 29.4) = -414,400\,\text{J/s}$$

this equals heat gained by cooling water

$$\dot{m}_{H_2O} = \left(\frac{160}{2}\right)\left(\frac{\pi(0.0166)^2}{4}\right)(997)(20) = 34.52\,\text{kg/s}$$

$$\dot{q} = \dot{m}C_p(T_{exit} - T_{inlet})$$

$$T_{H_2O,exit} = \dot{q}/\dot{m}C_p + T_{inlet} = 414,400/(34.52)(4180) + 24$$

$$= 26.9^\circ\text{C}$$

$$\Delta T_{\ln mean} = \frac{(33.9 - 26.9) - (29.4 - 24.0)}{\ln(33.9 - 26.9)/(29.4 - 24.0)} = 6.2^\circ\text{C}$$

Correct ΔT for the two tube pass arrangement by obtaining correction factor

$$P = \frac{26.9 - 24}{33.9 - 24} = 0.29 \qquad R = \frac{33.9 - 29.4}{26.9 - 24} = 1.55 \quad \text{From Fig.14-4} \quad F = 0.94$$

$$\Delta T = (0.94)(6.2) = 5.8^\circ\text{C}$$

An approximate value for the shell-side heat transfer coefficient in a cross-flow exchanger with segmented baffles and reasonable clearance between baffles, between tubes, and between baffles and shell can be obtained using the relation developed by Colburn.[1]

$$h_o = \frac{k}{D_o}\frac{a_o}{F_s}\left(\frac{D_o G_s}{\mu_f}\right)^{0.6}\left(\frac{C_p\mu}{k}\right)_f^{1/3} \quad \text{Subscript } f \text{ at average film temperature.}$$

---

[1] It can be estimated more accurately by using the Bell-Delaware method outlined in Example 14-6 or by using a computer software program.

14-10 (continued)

where $a_o = 0.33$ when tubes in a tube bank are staggered, F_s a correction factor to account for bypassing effects usually between 1.1 and 1.7, and G_s the shell-side mass velocity across the tube based on the minimum free area between baffles at the shell axis. The free area S for use in evaluating G_s for the case of a full-packed shell and transverse openings giving the smallest free area is estimated from

$$S = \frac{\text{(ID of shell)(clearance between adjacent tubes)(baffle spacing)}}{\text{center-to-center distance between adjacent tubes}}$$

$$= \frac{(0.387)(0.0238 - 0.0191)(0.3)}{0.0238} = 0.023\,\text{m}^2$$

$$G_s = 22/0.023 = 956.5 \text{ kg/m}^2 \cdot \text{s}$$

$$\text{Pr} = (C_p\mu/k)_f = (4179)(0.000786)/0.618 = 5.32 \qquad \text{at 304K film temperature}$$

$$\text{Re} = D_oG_s/\mu_f = (0.0191)(956.5)/0.000786 = 23{,}240$$

Assume an F_s of 1.3 to account for the bypassing effects.

$$h_o = (0.618/0.0191)(0.33/1.3)(23{,}240)^{0/6}(5.32)^{1/3} = 5973 \text{ W/m}^2 \cdot \text{K}$$

The heat transfer coefficient inside the tubes is given by Eq. (14-18). Assume negligible viscosity effects.

$$h_i = 0.023(k/D_i)\text{Re}^{0.8}\,\text{Pr}^{1/3}$$

$$\text{Re}_i = D_iV\rho/\mu = (0.0166)(2.0)(996)/0.000821 = 40{,}275$$

$$\text{Pr}_i = C_p\mu/k = (4179)(0.000821)/0.609 = 5.63$$

$$h_i = (0.023)(0.609/0.0166)(40{,}275)^{0.8}(5.63)^{1/3} = 7252 \text{ W/m}^2 \cdot \text{K}$$

The overall heat transfer coefficient is then obtained from Eq. (14-4). Assume tube is copper $k_w = 383 \text{ W/m} \cdot \text{K}$.

14-10 (continued)

$$\frac{1}{U_o} = \frac{D_o}{h_i D_i} + \frac{D_o}{h_{i,d} D_i} + \frac{D_o \ln(D_o / D_i) x_w}{k_w (D_o - D_i)} + \frac{1}{h_o} + \frac{1}{h_{o,d}}$$

$$= \frac{0.0191}{(7252)(0.0166)} + \frac{0.0191}{(5680)(0.0166)} + \frac{0.0191 \ln(0.0191/0.0166)}{383} + \frac{1}{5973} + \frac{1}{11{,}360}$$

$$= 6.213 \times 10^{-4}$$

$U_o = 1610\ \text{W/m}^2 \cdot \text{K}$

Area required is

$$A_o = \dot{q} / U_o \Delta T_{mean} = 414{,}400/(1610)(5.8) = 44.4\ \text{m}^2$$

Area available is

$$A_o = (160)\pi(0.0191)(4.876) = 46.8\ \text{m}^2$$

The area is adequate with copper tubes, but would not be adequate for steel tubes. †

Pressure drop calculation inside the tube using Eq. (14-23)

$$\Delta p_i = 2\beta_i + G_i^2 L n_p / \rho_i D_i \phi_i$$

Because of the small temperature change $\phi_i \cong 1$ and assuming few expansions and contractions because of the liquid flows, $\beta \cong 1$. For a Re of 40,275, $f = 0.0055$ from Fig.(12-1)

$$\Delta p_i = (2)(0.0055)[(2.0)(996)]^2(4.876)(2)/(996)(0.0166)(10^{-3}) = 25.75\ \text{kPa}$$

The inside pressure drop is adequate.

The shell-side pressure drop may be obtained with Eq.(14-24).

$$\Delta p_o = 2B_o f' N_{tr} G_s^2 / \rho_o$$

$$N_{tr} = 0.387/0.0238 = 16.3 \quad \text{as a maximum}$$

Combining Eqs.(14-24a) and 14-24c results in

---

† If the tubes were of steel, the U_o would be reduced to 1483 W/m²·K and the area required would be 48.2 m², slightly above that available.

14-10 (continued)

$$f' = \left[0.23 + \frac{0.11}{(X_T^{-1})^{1.08}}\right]\left(\frac{D_o G_s}{\mu_f}\right)^{-0.15}$$

$$= \left[0.23 + \frac{0.11}{\left(\frac{0.0238}{0.0191} - 1\right)^{1.08}}\right]\left(\frac{1}{23{,}240}\right)^{0.15} = 0.161$$

Assume that $B_o = 16$ as a maximum

$$\Delta p_o = (2)(16)(0.161)(16.3)(956.5)^2 / 995 = 77{,}220 \text{ Pa} = \mathbf{77.2 kPa}$$

ANSWER

The use of $B_o = 16$ and $N_{tr} = 16.3$ provides a large double safety factor in the calculation. The allowable pressure drop is 69 kPa, but the safety factor should be sufficient to accept the heat exchanger to cool 22 kg/s of distilled water.

Addendum:

This problem can be calculated using appropriate software such as CC-Therm. When this done, the overall heat transfer coefficient is about 10% higher and the pressure drop for the shell side is nearly 20% lower. Thus. the computer solution also indicates that the proposed heat exchanger can handle the specified heat load.

14-11

First calculate the ideal cross-flow heat transfer coefficient

$$\mathrm{Re} = D_o V_{\max} \rho / \mu \qquad \text{where } V_{\max} \text{ is given by Eq. (14-39) using Eq. (14-40)}$$

$$S_m = L_B[(D_s - D_{OTL}) + \frac{D_{OTL} - D_o}{P_T}(P_T - D_o)] \qquad D_{OTL} = D_s - \Delta_b = 0.54 - 0.035 = 0.505$$

$$= (0.127)[0.035 + \frac{0.505 - 0.0254}{0.03175}(0.03175 - 0.0254)] = 0.01662\,\mathrm{m}^2$$

$$V_{\max} = \frac{5.5}{(730)(0.01662)} = 0.4533\,\mathrm{m/s}$$

Use Eq. (14-22) to obtain the ideal heat transfer coefficient for a Reynolds number of 21,010 with a $\mu = 4\times10^{-4}\,\mathrm{Pa\cdot s}$

$$h_o = \frac{k}{D_o} a \mathrm{Re}^m \mathrm{Pr}^{0.34} F_1 F_2 \qquad \text{neglect the correction factors } F_1 \text{ and } F_2$$

$$= \left(\frac{0.13}{0.0254}\right)(0.211)\left[\left(2.1\times10^4\right)^{0.651}\right]\left[\frac{(2470(4\times10^{-4}}{0.13}\right]^{0.34} = 1402\,\mathrm{W/m^2\cdot K}$$

Now calculate correction factors J_c, J_L, and J_B in Eq. (14-42) and use Fig. 14-45

$$F_c = \frac{1}{\pi}\left[\pi + \frac{2(D_s - 2L_c)}{D_{OTL}}\sin\left(\cos^{-1}\frac{(D_s - 2L_c)}{D_{OTL}}\right) - 2\cos^{-1}\frac{(D_s - 2L_c)}{D_{OTL}}\right]$$

$$L_c = (0.25)D_s = (0.25)(0.54) = 0.135\,\mathrm{m}$$

$$\frac{D_s - 2L_c}{D_{OTL}} = \frac{0.54 - 2(0.135)}{0.505} = 0.5346$$

$$F_c = \frac{1}{\pi}\left[\pi + (2)(0.5346)\sin(\cos^{-1}0.5346) - 2\cos^{-1}0.5346\right]$$

$$= 0.647$$

$$J_c = 0.55 + (0.72)(0.647) = 1.016$$

To calculate J_C, determine the leakage areas from Eqs. (14-43a) and (14-43b)

14-11 (continued)

$$S_{sb} = D_s\left(\frac{\Delta_{sb}}{2}\right)\left[\pi - \cos^{-1}\left(1 - \frac{2L_c}{D_s}\right)\right]$$

$$= (0.54)\left(\frac{0.005}{2}\right)\left[\pi - \cos^{-1}\left(1 - \frac{2(0.135)}{0.54}\right)\right]$$

$$= 2.826\times10^{-3}\,\mathrm{m}^2$$

$$S_{tb} = \pi D_o\left(\frac{\Delta_{tb}}{2}\right)N_T\left(\frac{1+F_c}{2}\right)$$

$$= \pi(0.0254)\left(\frac{0.0008}{2}\right)(158)\left(\frac{1+0.647}{2}\right) = 4.153\times10^{-3}$$

$$\frac{S_{sb}+S_{tb}}{s_m} = \frac{2.826\times10^{-3}+4.153\times10^{-3}}{1.662\times10^{-2}} = 0.4199$$

$$\frac{S_{sb}}{S_{sb}+S_{tb}} = \frac{2.826\times10^{-3}}{2.826\times10^{-3}+4.153\times10^{-3}} = 0.4049$$

J_L is obtained from Fig. 14-46 as 0.57

To calculate J_B , obtain the bypass area fraction F_{bp} from Eq. (14-44)

$$F_{bp} = (L_B / S_m)(D_s - D_{OTL}) = (0.127/1.662\times10^{-2})(0.035) = 0.2674$$

For $N_{ss}/N_c = 0.2$, Fig. 14-47 provides a value for J_B of 0.92

The individual heat transfer coefficient on the shell side is then

$$h_s = h_o J_C J_L J_B = (1402)(1.016)(0.57)(0.92) = \underline{\underline{747\ \mathrm{W/m^2\cdot K}}}$$

Answer

The procedure to obtain the pressure drop on the shell side follows the steps outlined in Eqs. (14-46) through (14-51). The ideal cross-flow pressure drop is obtained from

14-11 (continued)

$$\Delta p_c = (K_a + N_c K_f)(\rho V_{\max}^2 / 2) \quad \text{neglect } K_a \text{, obtain } K_f \text{ from Table 14-10}$$

$$K_f = 0.267 + \frac{0.249\times10^4}{21{,}010} - \frac{0.927\times10^7}{(21{,}010)^2} + \frac{0.1\times10^4}{(21{,}010)^3} = 0.3625 \text{ for } 2\times10^4 < \text{Re} < 2\times10^6$$

$$N_c = D_s(1 - 2L_c / D_s) / P_{TD}$$

$$= (0.54)\left[1 - 2(0.135)/0.54\right]/(0.03175) = 8.5$$

The cross-flow pressure drop on one baffle space is obtained from

$$\Delta p_c = (8.5)(0.3625)(730)(0.4533)^2 / 2 = 231\,\text{Pa}$$

Now calculate the pressure drop for the window zone between baffles with the window flow area given by Eq. (14-49)

$$S_w = (D_s^2/4)\left[\cos^{-1} D_B - D_B(1 - D_B^2)^{1/2}\right] - (N_T/8)(1 - F_c)\pi D_o^2$$

$$D_B = (D_s - 2L_s)/D_s = [0.54 - 2(0.135)]/0.54 = 0.5$$

$$S_w = (0.54)^2/4\left\{\cos^{-1} 0.5 - 0.5[1 - (0.5)^2]^{1/2}\right\} - (158/8)(1 - 0.647)\pi(0.0254)^2$$

$$= 0.04474 - 0.01414 = 0.0306\,\text{m}^2$$

Calculate the effective number of cross flow rows in the window zone

$$N_{cw} = 0.8L_c / P_{TP} = (0.8)(0.135)/0.03175 = 3.40$$

The window zone pressure drop is given by Eq. (14-47b)

$$\Delta p_w = (2 + 0.6N_{cw})\dot{m}_T^2 / 2S_m S_w \rho \quad \text{where } \dot{m}_T = 5.5\,\text{kg/s}$$

$$= [2 + 0.6(3.4)](5.5)^2/(2)(0.01662)(0.0306)(730) = 164.6\,\text{Pa}$$

Now use Figs. 14-48 and 14-49 to obtain leakage and bypass correction factors R_B and R_L for use in Eq. (14-51).

From Fig. 14-48, $R_B = 0.76$

Fig. 14-49, $R_L = 0.33$

$$\Delta p_s = [(N_B - 1)\Delta p_c R_B + N_B \Delta p_w]R_L + 2\Delta p_c R_B (1 f N_{cw} / N_c)$$

where N_B from Eq. (14-36) is $L_s/(L_B + t_b) - 1$

14-11 (continued)

$$N_B = \frac{4.267}{0.127 + 0.005} - 1 = 31.3 \text{ baffles}$$

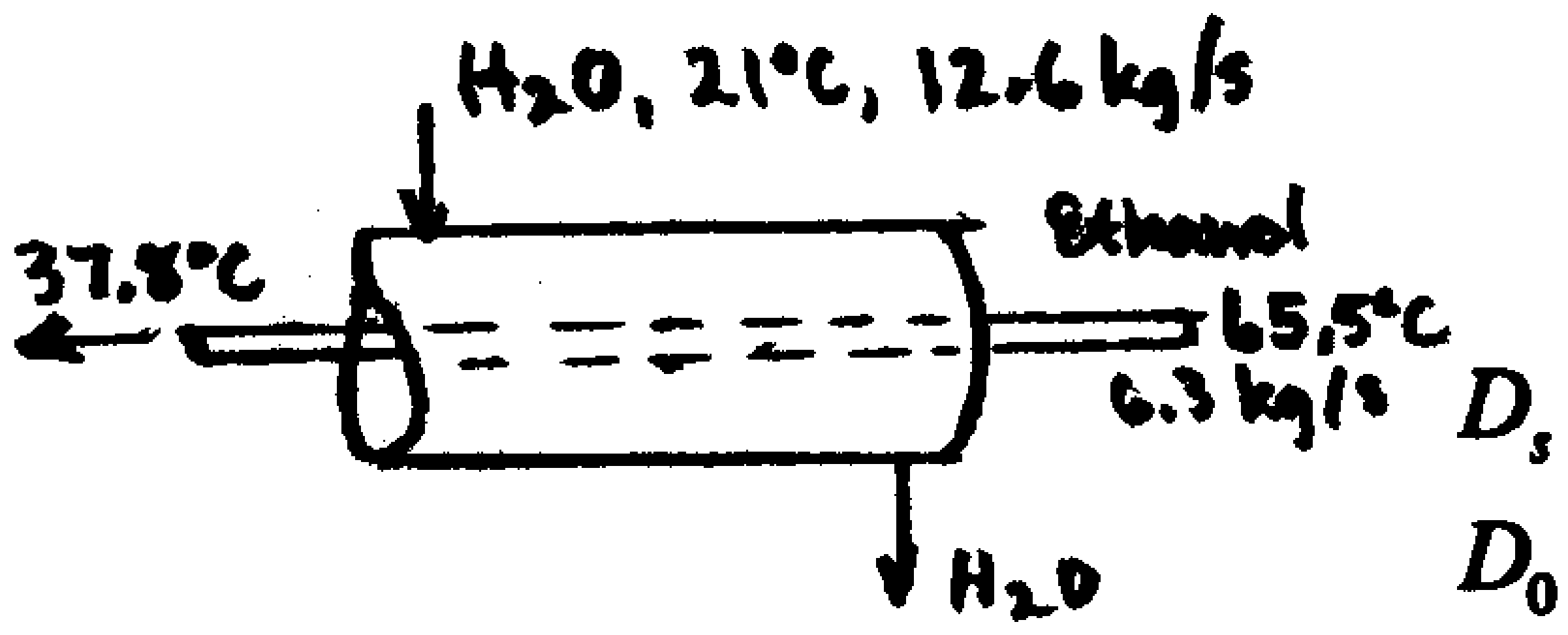
$$\text{or } 31$$

$$\Delta p_s = [(31-1)(231)(0.76) + (31)(165)](0.33) + 2(231)(0.76)(1 + 3.4/8.5)$$

$$= 3426 + 492 = \underline{\underline{\mathbf{3918\ Pa\ \ or\ \ 3.9\ kPa}}}$$

Answer

14-12



$D_s = 0.127\text{m}$ $N_T = 7$

$D_0 = 0.0254\text{m}$

$D_i = 0.0203\text{m}$

For the alcohol,

$T_{avg} = 51.65°\text{C}$

$C_p = 2512 \text{J/kg °C}$

$\rho = 770 \text{kg/m}^3$

$k = 0.151 \text{ W/m} \cdot \text{K}$ (Appendix D)

$$\dot{q}_{alc} = \dot{m}C_p\Delta T = (6.3)(2512)(65.5-37.8)(10^{-3})$$
$$= 438.4 \text{kJ/s}$$
$$\Delta T_{H_2O} = \dot{q}/\dot{m}C_p = (438{,}400)/(12.6)(4186)$$
$$= 8.3°\text{C} \quad T_{ave,H_2O} = 25.1°\text{C}$$

Velocity of ethyl alcohol

$$V_{alc} = (6.3 \text{kg/s}/7)(\frac{1}{770 \text{kg/m}^3})\frac{4}{\pi(0.0203)^2 \text{m}^2} = 3.61 \text{m/s}$$
$$\text{Re} = DV\rho/\mu = (0.0203)(3.61)(770)/0.0007 = 80{,}600$$

Calculate h_{alc} with Eq.(14-18) assuming $(\mu/\mu_w)^{0.14} \cong 1$

$$h_{alc} = (0.151/0.0203)(0.0203)(80{,}600)^{0.8}[(2512)(0.0007)/0.151]^{1/3}$$
$$= 3263 \text{ W/m}^2 \cdot \text{K}$$

For the water side, the available cross-sectional area available for flow is

$$A = \frac{\pi D_S^2}{4} - \frac{7\pi D_0^2}{4} = \frac{\pi(0.127)^2}{4} - \frac{7\pi(0.0254)^2}{4} = 0.00915 \text{m}^2$$
$$V_{H_2O} = (12.6)/(996)(0.00915) = 1.38 \text{m/s}$$

The combined diameter for the water side is estimated by assuming that the 7 tubes are combined into one tube or $7\pi D_0^2/4 = \pi D_{comb}^2/4$.

14-12 (continued)

$$D_{combined} = \left[7(0.0254)^2\right]^{1/2} = 0.0672\text{m}$$

as suggested by the <u>*Heat Exchanger Design Handbook*</u>, the effective diameter for the water side can be taken as 0.127-0.0672 = 0.0598m.

$$\text{Re} = D_e \rho V / \mu = (0.0598)(996)(1.38)/0.000899 = 91,400$$

$$\text{assume } h_{H_2O} = \left(\frac{0.610}{0.0598}\right)(0.023)(91,400)^{0.8}\left[(4180)(0.000899)/0.610\right]^{1/3}$$
$$= 4002\ \text{W/m}^2\cdot\text{K}$$

Base overall coefficient on outside area of tubes

$$\frac{1}{U_0} = \frac{1}{4002} + \frac{0.0254}{(3263)(0.0203)} + \frac{(0.0254)(0.00254)}{(45)(0.02286)}$$
$$= 2.4487\times10^{-4} + 3.8346\times10^{-4} + 6.272\times10^{-5}$$
$$= 6.9105\times10^{-4}$$
$$U_0 = 1447\ \text{W/m}^2\cdot\text{K}$$

$$\Delta T_{\ln mean} = \frac{(65.5-29.3)-(37.8-21)}{\ln 36.2/16.8} = \frac{19.4}{0.767} = 25.3^\circ\text{C}$$

Outside tube area is then

$$A = \dot{q}/U_0 \Delta T_{\ln m} = 438,400/(1447)(25.3) = 12.0\,\text{m}^2$$

Tube length

$$L = (12.0/7)/(\pi)(0.0254) = 21.5\,\text{m}$$

From the mechanical energy balance, Eq.(12-2), the work that must be provided in the pumping process reduces to

$$W = \sum F = 2fV^2 L/D$$

From Fig. 12-1 values for the friction factor for smooth pipes are

14-12 (continued)

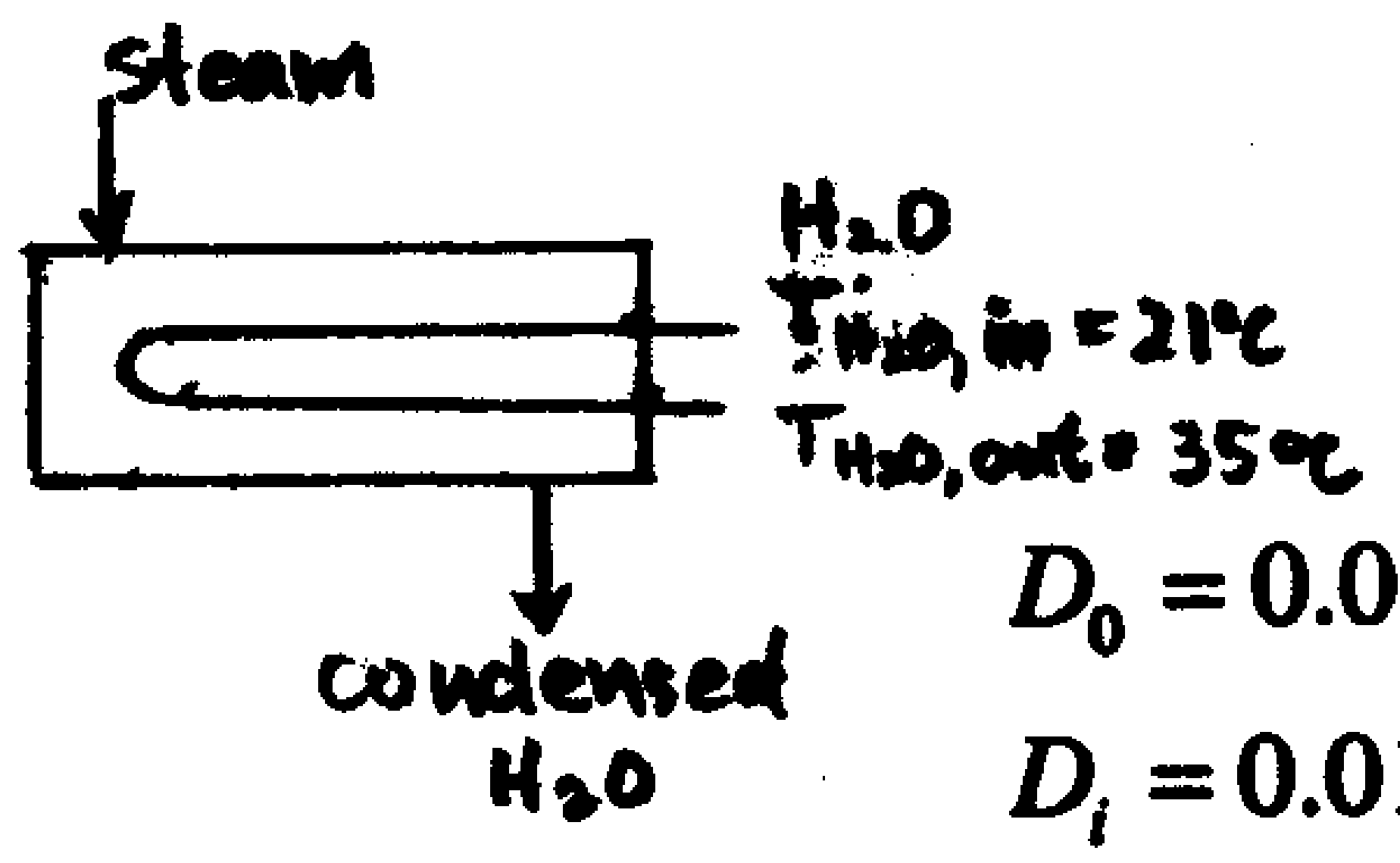
$$\mathrm{Re} = 80{,}600 \quad f = 0.0061$$
$$\mathrm{Re} = 91{,}400 \quad f = 0.0059$$
$$\sum F_{alc} = (2)(0.0061)(3.61)^2(21.5)(1.2)/0.0203 = 202\ \mathrm{Nm/kg}$$
$$\sum F_{H_2O} = (2)(0.0059)(1.38)^2(21.5)(1.2)/0.0598 = 9.7\ \mathrm{Nm/kg}$$

Power cost

$$\mathrm{Cost}_{alc} = \left(\frac{202}{0.6}\right)(6.3)\left(\frac{1}{1000}\right)(300)(24)(0.08) = \$1222/\mathrm{y}$$
$$\mathrm{Cost}_{H_2O} = \left(\frac{9.7}{0.6}\right)(12.6)\left(\frac{1}{1000}\right)(300)(24)(0.08) = \$117/\mathrm{y}$$

Total pumping cost = 1222 + 117 = **$ 1339/y**
Answer

14-13



$D_0 = 0.0159\text{m}$ 18 gauge $L_p = 1.83\text{m}$ $T_{sat} 51.7°\text{C}$

$D_i = 0.0134\text{m}$ 129 tubes/pass $V_{H_2O} = 1.22\text{m/s}$

Total area for heat transfer based on D_0

$$A = (\pi D_0)L_p N_T(2) = \pi(0.0159)(1.83)(129)(2) = 23.6\text{m}^2$$

Total cross-sectional area for water flow/tube pass

$$A_C = (\pi D_i^2/4)N_T = (\pi/4)(0.0134)^2(129) = 0.0182\text{m}^2$$

Although this is a tube-tube pass heat exchanger, the mean temperature difference is equal to the log mean temperature difference with the correction factor F = 1

$$\Delta T_{lnm} = \frac{(51.7-21)-(51.7-35)}{\ln 30.7/16.7} = 23°\text{C}$$

To simplify the problem evaluate the overall heat transfer coefficient at the location where the water temperature is (21+35)/2=28°C. To obtain the heat transfer coefficient by Eq.(14-78) requires a wall temperature. Since the thermal resistance in the water stream is at least twice that in the condensing steam, assume a wall temperature of 46°C. At a fluid temperature of 28°C and a film temperature of 37°C, the properties for water are

$\rho = 996\text{kg/m}^2$ $\mu = 0.838\,\text{cP}$ $C_p = 4179\text{J/kg}\cdot\text{K}$

$k = 0.614\text{W/mk}$ $\mu_w = 0.710\,\text{cP}$

$$h_i = (k/D)(0.023)(D_i V\rho/\mu)^{0.8}(C_p\mu/k)^{1/3}(\mu/\mu_w)^{0.14}$$

$$= \left(\frac{0.614}{0.0134}\right)(0.023)\left[\frac{(0.0134)(1.22)(996)}{8.38\times10^{-4}}\right]^{0.8}\left[\frac{(4179)(8.38\times10^{-4})}{0.614}\right]^{1/3}\left[\frac{8.38\times10^{-4}}{7.10\times10^{-4}}\right]^{0.14}$$

$$= 5195\text{W/m}^2\cdot\text{K}$$

For the condensing steam use Eq.(14-78) for a film temperature of 48.9°C.

14-13 (continued)

$$\bar{h}_0 = 0.725\left[\frac{k_L^3\rho_L(\rho_L-\rho_V)g\lambda_c}{\mu_L D_0(T_{sat}-T_w)}\right]^{1/4} \qquad \lambda_c = 2388\,\mathrm{kJ/kg}$$

$$\bar{h}_0 = 0.725\left[\frac{(0.642)^3(989)(989-0.07)(9.806)(2388\times10^3)}{(0.000567)(0.0159)(51.7-46)}\right]^{1/4}$$

$$= 13{,}435\,\mathrm{W/m^2\cdot K}$$

$$\frac{1}{U_0} = \frac{0.0159}{(5195)(0.0134)} + \frac{(0.0159)\ln(0.0159/0.0134)}{(2)(99.7)} + \frac{1}{12{,}450}$$

$$U_0 = 3160\,\mathrm{W/m^2\cdot K}$$

Check on the wall temperature

$$(3160)(51.6-28) = 13{,}435(51.7-T_w)$$

$T_w = 46.1^\circ C$ compared to an assumed wall temperature of $46^\circ\mathrm{C}$; further iteration with a new wall temperature is not necessary. Thus the overall heat transfer coefficient is $3160\,\mathrm{W/m^2\cdot K}$

Answer

The kilograms of steam condensed is

$$\dot{q}/\lambda = UA\Delta T_{\ln m}/\lambda$$

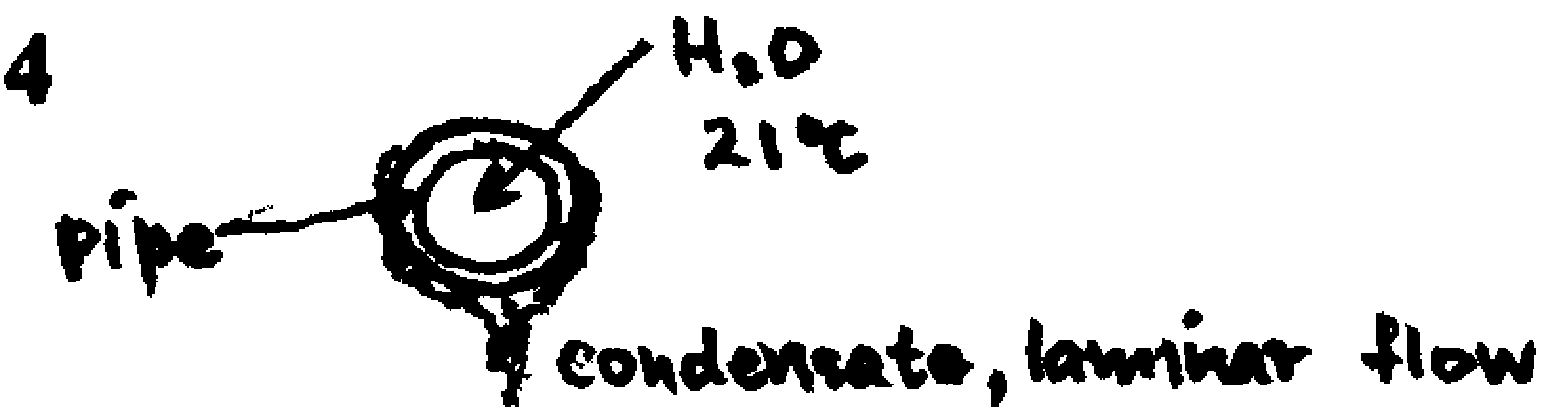
$$= (3160)(23.6)(23)/2388\times10^3$$

$$= 0.72\,\mathrm{kg/s}$$

Answer

Note that the calculation has not included various operating conditions that could alter the condensation process such as fouling effects, presence of noncondensibles in the steam, subcooling of the condensate, etc.

14-14



$D_0 = 0.030\text{m}$

$D_i = 0.025\text{m}$

Determine h_i from Eq.(14-18). Assume a velocity of 1 m/s for the water flow

$$\text{Re}_{H_2O} = (DV\rho/\mu)_i = (0.025)(1)(998)/0.000982 = 25,400$$

$$\text{Pr}_{H_2O} = \left(\frac{C_p\mu}{k}\right) = (4186)(0.000982)/0.599 = 6.86$$

$$h_{H_2O} = \left(\frac{0.599}{0.025}\right)(0.023)(25,400)^{0.8}(6.86)^{1/3} \quad \text{assume}\left(\mu/\mu_w\right)^{0.14} \cong 1$$

$$h_{H_2O} = 3498\ \text{W/m}^2\cdot\text{K}$$

Calculations for the outside condensation coefficient uses Eq. (14-78). However, since the temperature of the wall T_w is unknown, this requires an iterative calculation that is best performed with a simple computer software program. The procedure is as follows:

> Assume a T_w and calculate the condensation coefficient. This permits evaluation of an overall coefficient and a heat transfer rate. From the latter T_w may be evaluated. If the value does not match the initial assumption, modify T_w and repeat the process. The first step is provided below.
>
> The average T_w would be $(127+21)/2 = 74^\circ\text{C}$ or 347K. Begin with a T_w of 330 K.

$$h_{i-C_S} = 0.725\left[\frac{k_L^3\rho_L(\rho_L-\rho_v)g\lambda_C}{\mu_L D_0(T_{sat}-T_w)}\right]^{1/4}$$

$$= 0.725\left[\frac{(0.082)^3 551(551-14.1)(9.806)(283,800)}{(0.0001285)(0.03)(400-330)}\right]^{1/4}$$

$$= 825\ \text{W/m}^2\cdot\text{K}$$

$$\frac{1}{U_0} = \frac{1}{825} + \frac{0.03}{(3498)(0.025)} = 1.5552\times10^{-3}$$

$$U_0 = 643$$

$$\dot{q} = -U_o(T_{sat}-T_w) = 643(400-294) = 68,160\ \text{W/m}^2$$

$$T_w = T_{sat} - \dot{q}/h_{i-C_S} = 400 - 68,160/825 = 317.4\text{K}$$

Temperature assumed is high and needs to be lowered for the next iterarion.

14-14 (continued)

For an assumed T_w of 317.4K

$$h_{i-C_5} = 792\,\mathrm{W/m^2 \cdot K}$$
$$U_0 = 622.6\,\mathrm{W/m^2 \cdot K}$$
$$T_w = 316.7\,\mathrm{K}$$

after several iterations the final values are

$$U_0 = \underset{\textbf{Answer}}{\underline{\underline{\mathbf{619 W/m^2 \cdot K}}}} \qquad \dot{q} = \underset{\textbf{Answer}}{\underline{\underline{\mathbf{66.2 kW/m^2}}}}$$

The rate of condensation per unit length of tube is

$$\begin{aligned}\dot{m}_{cond} &= (\dot{q}/\Delta H_c)\pi D \\ &= (66.2/283.8)\pi(0.03) = \underset{\textbf{Answer}}{\underline{\underline{\mathbf{0.022 (kg/s)/m}}}}\end{aligned}$$

14-15

The CC-Therm software is used to design the partial condenser E-601. The tabulated data and results are shown below.

	Tube-side	Shell-side
Flow, kg/s	35.30	64.57
Non-condensed flow, kg/s	-	17.07
Operating pressure kPa	109	194.4
Press-drop allowed, kPa	34.5	34.5
Outside tube diameter, m	0.0127	-
Inside tube diameter, m	0.0094	-
Length of tubes, m	3.05	-
Temperature of fluids, °C	30.1	-46
Heat exchanged, kW	1565.3	-1565.3

Results:

	Tube-side	Shell-side
Re	8091	952
Velocity, m/s	2.05	2.89
h (clean), $W/m^2 \cdot K$	3161	5838
Δp (calc), kPa	5.7	5.65

$\Delta T_{lnm} = 16.2°C$

$U(\text{calc})^* = \underline{\mathbf{807 W/m^2 \cdot K}}$

Answer

$A_{required}(\text{calc})^+ = \underline{\underline{\mathbf{1198 m^2}}}$

Answer

$N_{T\,required} = \underline{\underline{\mathbf{2688\ Tubes}}}$

Answer

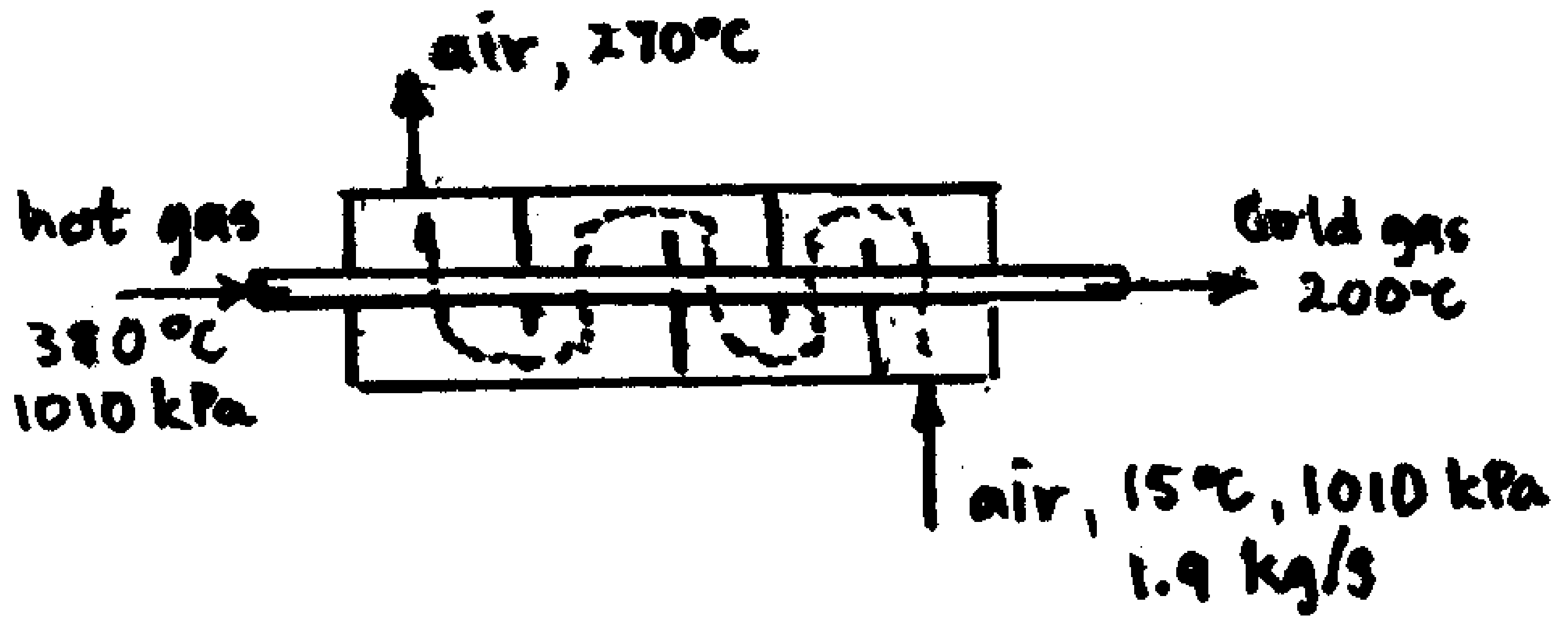
Inside shell diameter = $\underline{\underline{\mathbf{1.32\ m}}}$

Answer

* This includes a fouling factor of $1.76 \times 10^{-4} m^2 \cdot K/W$ for both sides of the tube.

\+ To meet the area requirement requires some ineffective area that is required to mechanically assemble the shell-and-tube heat exchanges. For example, for a class R, type AEL, the actual area needs to be increased to 1211 m^2.

14-16



Heat exchanger, sq. pitch, 0.0381 m
$D_i = 0.0191$ m
$D_o = 0.0254$ m
5000 h/y
Cost data use Fig. 14-19; power cost is \$0.12/kWh

The optimum length of the heat exchanger can be estimated with a hand calculation similar to the one that was used in the solution of Example 5 in Chapter 15 of the 4th edition of *Plant Design and Economics for Chemical Engineers.* The optimum length can also be estimated by using available computer software and rapidly examining a number of lengths, determining the pressure drop of the shell and tube side of the exchanger and the required area to establish an annual operating and fixed cost to identify the exchanger requiring the minimum annual cost. The latter approach has been taken to solve this problem.

The procedure is similar to that outlined in Example 14-9. Based on average temperatures on both sides of the exchanger, an estimate of the heat load can be obtained. With an estimated overall heat transfer coefficient, a preliminary area can be established to minimize the number of heat exchanger lengths that will need to be examined. With a heat load of approximately 530 kJ/s, it will require a heat exchanger of at least several meters to meet the heat load. Thus, the computer software was used to examine exchanger lengths above 4m. Near 6.1 m, a minimum in the annual operating and fixed cost was established. It should be noted that the cost curve was rather flat near the minimum.

Length of exchanger = **6.096 m**

Answer

Purchased cost for the optimum heater = **\$11,200**

Answer

15:1

15-1

A material balance for the depropanizer follows the procedure outlined in Example 15-2. The results are shown below.

	Mol %			Mols/100 kg mol of feed	
	Feed	Distillate	Bottoms	Distillate	Bottoms
Methane	26.0	43.5		26.0	
Ethane	9.0	15.0		9.0	
Propane	25.0	41.0	1.0	24.6	0.4
n-Butane	17.0	0.5	41.7	0.3	16.7
n-Pentane	11.0		27.4		11.0
n-Hexane	12.0		29.9		12.0
	100.0	100.0	100.0	59.9	40.1

Determine the minimum stages with Eq. (15-1), assume 96°C as the average temperature for the distillation column

$$N_{\min} = \ln\left[(x_{LK}/x_{HK})_D(x_{HK}/x_{LK})_B\right]/\ln(\alpha_{LK/HK})$$

$$= \ln\left[(0.41/0.005)(0.417)/0.01)\right]/\ln(1.55/0.8)$$

$$= \mathbf{12.3\,stages}$$

Answer

Use the Underwood method to calculate $R_{\min}$ using Eqs. (15-2) and (15-3)

$$\sum_{i=1}^{n}\frac{\alpha_i x_{F,i}}{\alpha_i - \Theta} = 1 - q \qquad \text{where } q = 0.34$$

$$\frac{18.75(0.26)}{18.75-\Theta}+\frac{4.75(0.09)}{4.75-\Theta}+\frac{1.9375(0.25)}{1.9375-\Theta}+\frac{1(0.17)}{1-\Theta}+\frac{0.48(0.11)}{0.48-\Theta}+\frac{0.24(0.12)}{0.24-\Theta}=0.66$$

After several iterations, $\Theta = 1.351$

$$R_{\min} + 1 = \sum_{i=1}^{n}\alpha_i x_{D,i}/(\alpha_i - \Theta)$$

$$R_{\min} = \frac{18.75(0.435)}{18.75-1.351}+\frac{4.75(0.15)}{4.75-1.351}+\frac{1.9375(0.41)}{1.9375-1.351}+\frac{1(0.005)}{1-1.351}-1$$

$$= 0.4687+0.2096+1.3544-0.0142-1$$

15-1 (continued)

$= 1.0185$ or **1.02**

Answer

Determine number of theoretical stages with the use of Eq. (15-6) with $R = 1.5(1.02) = 1.53$

$$\frac{N - N_{\min}}{N+1} = 0.75\left[1 - \left(\frac{R - R_{\min}}{R+1}\right)^{0.566}\right]$$

$$\frac{N - 12.3}{N+1} = 0.75\left[1 - \left(\frac{1.53 - 1.02}{1.53+1}\right)^{0.566}\right]$$

Solving for N results in **N=23 theoretical stages**.

Answer

The location of the feed is obtained from Eq. (15-6)

$$\log\frac{N_D}{N_B} = 0.206\log\left\{\left(\frac{B}{D}\right)\left(\frac{x_{HK}}{x_{LK}}\right)_F\left[\frac{(x_{LK})_B}{(x_{HK})_B}\right]^2\right\}$$

$$= 0.206\log\left\{\left(\frac{.401}{.599}\right)\left(\frac{0.17}{0.25}\right)\left(\frac{0.01}{0.005}\right)^2\right\}$$

$$N = N_D + N_B$$

Solving these two equations shows that the feed is located **13 stages** above the reboiler.

Answer

Use of a computer simulation provides the following results:

N_{min} (minimum stages) = 13 stages

R_{min} (minimum reflux) = 1.068

N (theoretical stages) = 25 stages

N_B (feed location above reboiler) = 15 stages

15:3

15-2

A material balance around the C_2 splitter is similar to that developed in Problem 3-6. The distillate shows a concentration of 99.9 mol percent ethylene, 0.093 mol percent ethane, and 0.007 mol percent methane. Likewise, the bottoms then provide a 99.43 mol percent ethane, 0.56 mol percent ethylene, and 0.01 mol percent propylene. Essentially this is a two component distillation similar to that presented in Examples 15-2 and 15-3. The material balance in kg/h is

Component†	Feed (601)	Distillate (602)	Bottoms (603)
Methane	2.9	2.9	--
Ethylene	62,121	62,067	62.9
Ethane	11,974	62.1	11,903
Propylene	2.1	--	2.1
	74,100	62,132	11,968

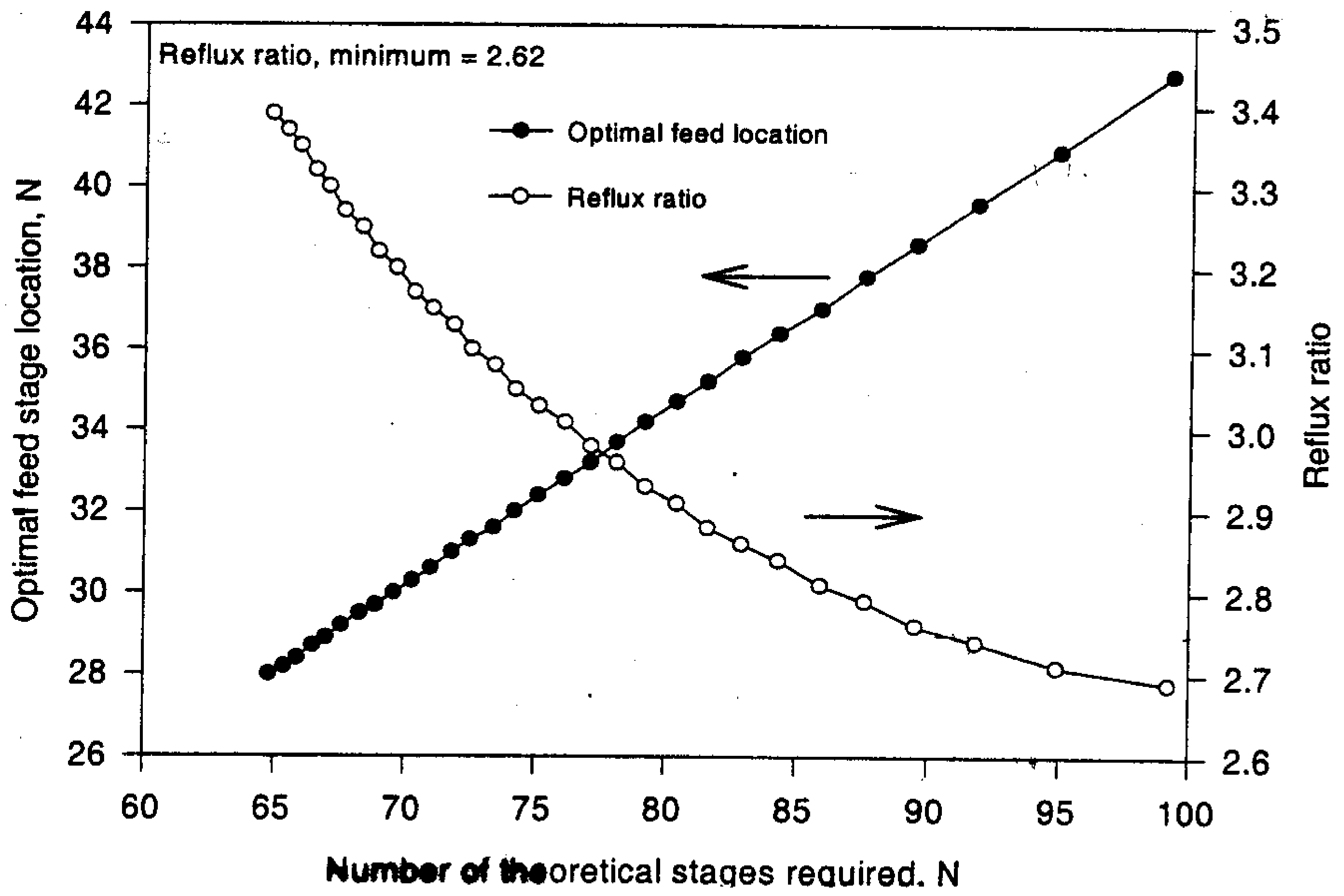
† Acetylene and propane have been neglected in the material balance.

The minimum reflux ratio can be evaluated for this two component distillation by using the Fenske-Underwood-Gilliland method and then determining what ratio factor to use to obtain the desired separation using 94 theoretical trays. This approach uses Eq. (15-1), (15-2), (15-3) and (15-4). If this approach is used, $N_{min} = 21.2$ stages and $R_{min} = 2.62$. A trial and error calculation with Eq. (15-4) where R is unknown, establishes that a value of 2.75 for R is required to obtain 94 theoretical trays. Thus $R = (1.05)(2.62)$ or 2.75 for this column. This is reasonable since the ratio factor for low temperatures distillation columns is generally between 1.05 and 1.10.

Answer

With the aid of a suitable computer software program, it is possible to investigate the effect of a number of reflux ratios on the number of stages that will be required to obtain the desired separation between ethylene and ethane. This approach is outlined on the computer software output which plots the effect of decreased reflux on the number of theoretical plates required for the separation.

15-2 (continued)



Note that for 94 theoretical plates a reflux ratio of 2.75 is required to make the desired separation. Since R_{min} was also established as 2.62 by the computer software, the R_{min} was also increased by a factor of 1.05.

The computer software also established the feed location for a 94 theoretical plate column as the **41st plate** above the reboiler.

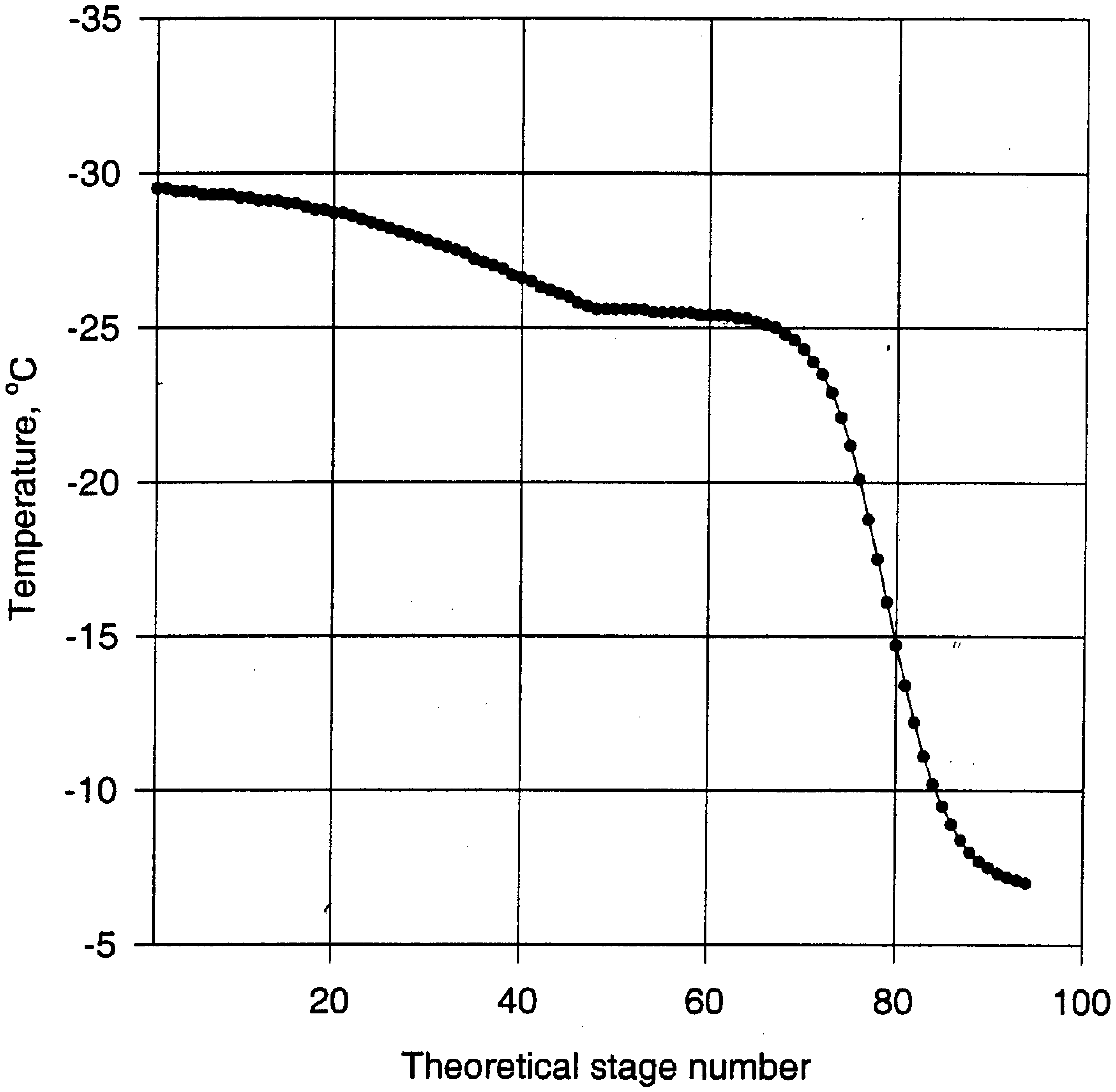
Answer

The temperature profile for the column examined in this problem is shown on the attached figure.

Answer

15:5

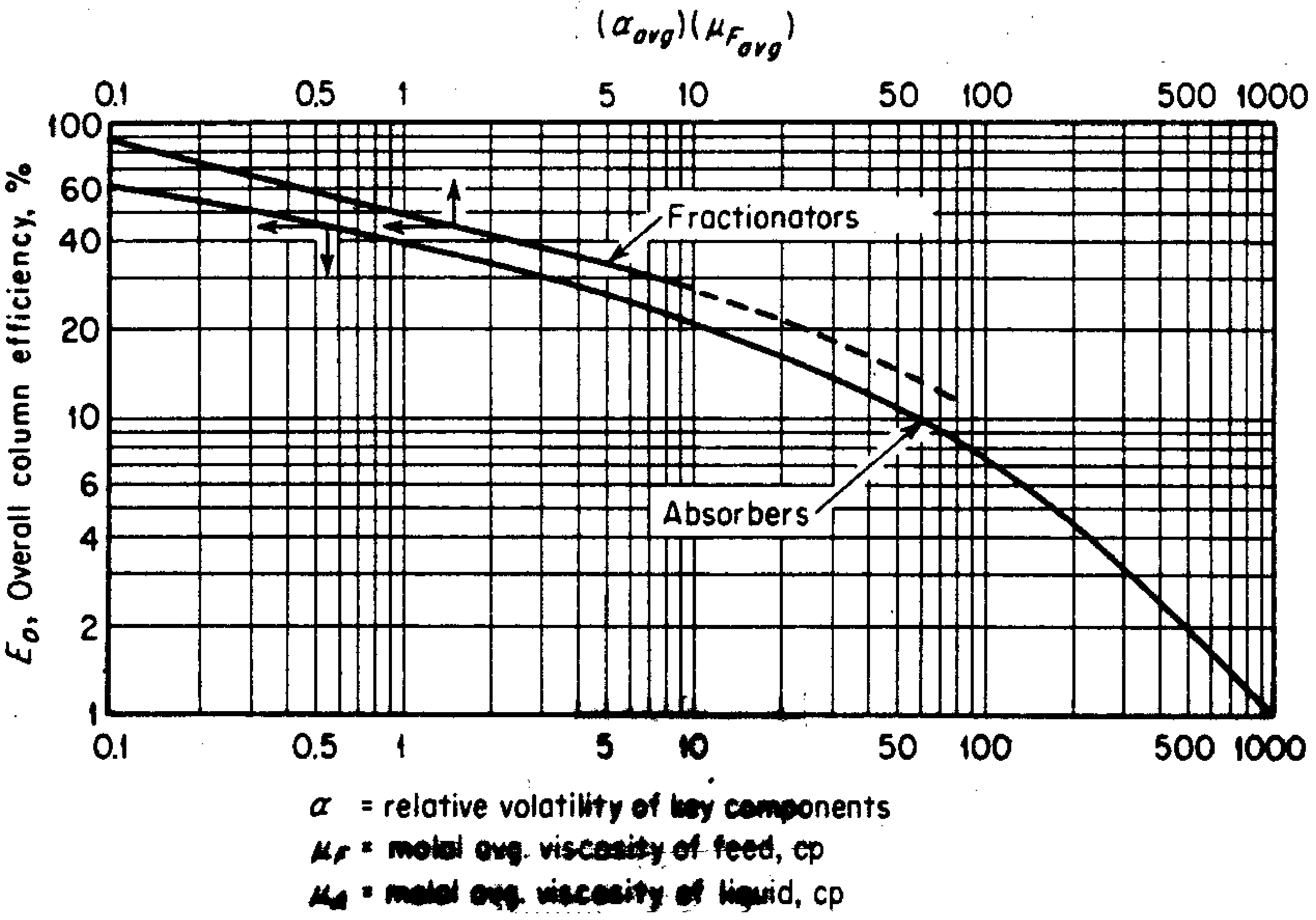
15- 2 (continued)



Temperature profile for the ethylene-ethane column as a function of theoretical stage number.

15-3

The overall column efficiency for finite-stage fractinators and absorbers has been correlated by O'Connell[†] and is shown in the figure below.



Overall column efficiencies for finite-stage contactor fractionators and absorbers.

To determine the overall column efficiency E_o, determine $\alpha_{\text{avg}}\mu_{\text{avg}}$

$T_{\text{avg}} = (82.8 + 100.5)/2 = 91.6^\circ\text{C}$ $\qquad$ $\alpha_{\text{avg}} = 1070/429 = 2.49$

At 91.6°C $\mu_{\text{Bz}} = 0.28\text{cP}$ $\qquad$ $\mu_{\text{Tol}} = 0.30\text{cP}$ $\qquad$ from Appendix D

$\mu_{\text{avg}} = \mu_{\text{molal avg}} = (0.28)(0.6) + (0.30)(0.4) = 0.288\,\text{cP}$

$E_o = 52\%$ from above figure

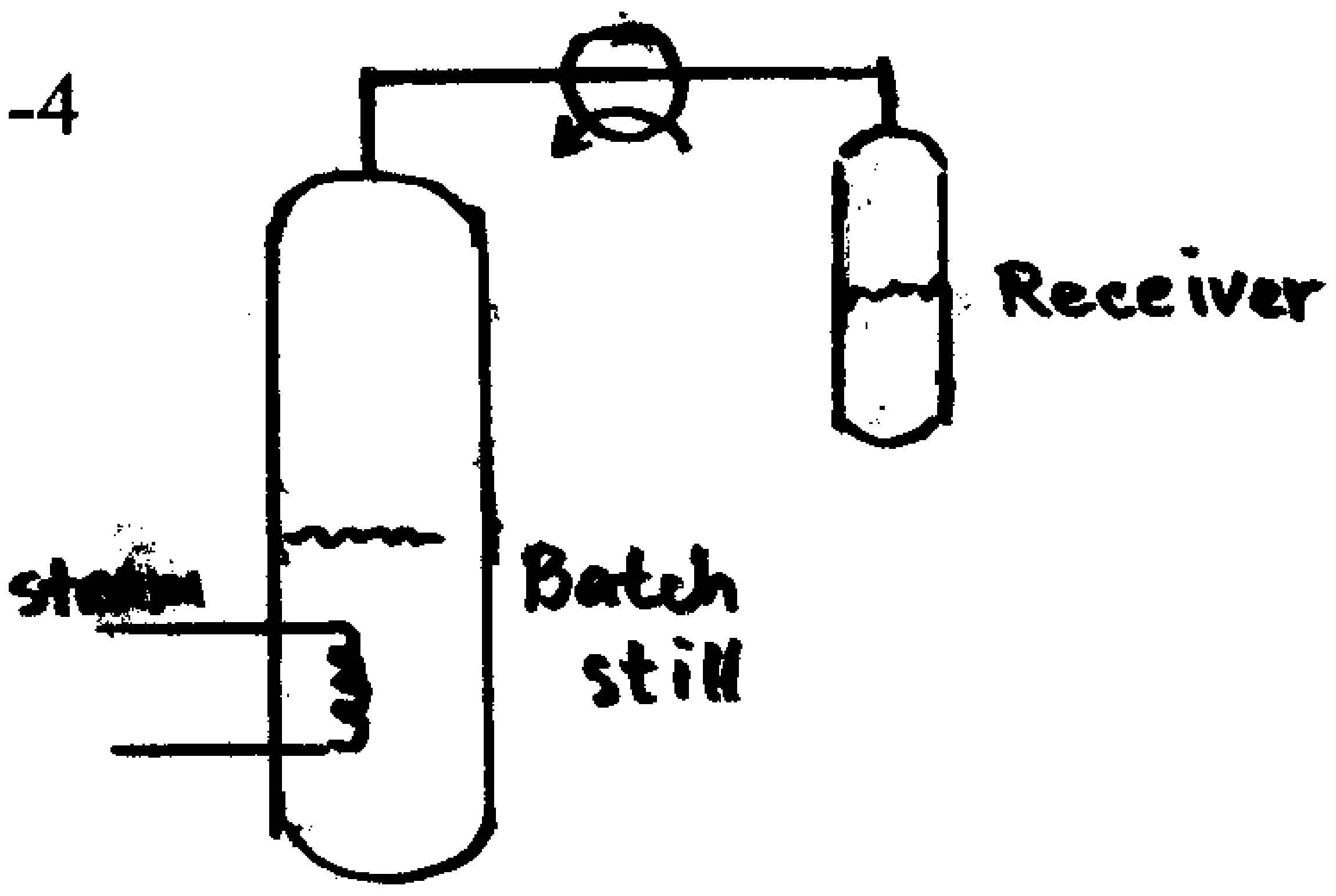
Assume reboiler equivalent to 1 theoretical stage

$N_{\text{act}} = (6.1-1)/0.52 = 9.8$ or **<u>10 actual trays</u>**

Answer

[†] H.E. O'Connell, *Trans. AICHE* **42**, 751 (1946).

15-4



Separate a mixture of 60 mol % benzene and 40 mol % toluene in a batch distillation to obtain a residual of 10 mol % benzene.

Assume α is constant at 2.9 and column holdup is negligible.

Under the above assumptions, the Rayleigh equation can be integrated, resulting in Eq. (15-29).

$$\ln\frac{L_2'}{L_2'} = \frac{1}{\alpha_{ij}-1}\ln\frac{x_{i2}(1-x_{i1})}{x_{i1}(1-x_{i2})} + \ln\frac{1-x_{i1}}{1-x_{i2}}$$

Substituting, assume $L_1' = 100\,\text{kg mols}$

$$= \frac{1}{1.9}\ln\frac{0.1(1-0.6)}{0.6(1-0.1)} + \ln\frac{(1-0.6)}{(1-0.1)}$$

$$= -1.3698 - 0.8109 = -2.1807$$

$$L_2' = (100)(0.113) = 11.3\,\text{kg mols}$$

By a material balance

	Batch Still		
	Start	End	Receiver
Benzene	60	1.13	58.87
Toluene	40	10.17	29.83
	100	11.3	88.7

Mol fraction of benzene in the receiver is

Mol fraction = 58.87 / 88.7 = **0.664 or 66.4%**

Answer

15-5

The required column diameter and pressure drop in a packed absorption column can be estimated quickly with the use of Fig. 15-7 or obtained quite accurately with one of several appropriate computer simulation programs. Both approaches will be used in the solution of this problem.

To use Fig. 15-7 requires developing a value for the abscissa and from this obtaining the superficial gas mass flux G under flooding conditions and from this the maximum allowable value of G. From this both the column diameter and pressure drop can be determined.

$V = 900/A_c$ $L = 2700/A_c$ where A_c is the cross-sectional area

$$\frac{L}{V}\left[\rho_V/(\rho_L-\rho_V)\right]^{1/2} = (2700/900)\left[1.2(881-1.2)\right]^{1/2} = 0.1108$$

At an abscissa value of 0.1108 on Fig. 15-7

$$G^2F_p\mu_L^{0.1}/\rho_v(\rho_L-\rho_v) = 0.18 \qquad F_p = 50 \text{ from Fig. 15-7}$$

Thus

$$G_{\text{flood}} = \left[\frac{\rho_v(\rho_2-\rho_v)(0.18)(3600)^2}{F_p\mu_L^{0.1}}\right]^{1/2} = \left[\frac{(1.2)(881-1.2)(0.18)(3600)^2}{(50)(0.02)^{0.1}}\right]^{1/2}$$

$$= 8535 \text{ kg/h}\cdot\text{m}^2$$

$$G_{\text{allowable}} = (8535)(0.7) = 5975\,\text{kg/h}\cdot\text{m}^2$$

This is equivalent to an ordinate value of

$$\frac{(5975/3600)^2(50)(0.02)^{0.1}}{(1.2)(881-1.2)} = 0.088$$

For a value of 0.088 $\underline{\Delta p \cong \mathbf{700\,Pa}/\text{meter}}$

Answer

$$\underline{D = (A_c/0.785)^{1/2} = \left[(900/5975)/0.785\right]^{1/2} = \mathbf{0.44\,m}}$$

Answer

With a computer simulation program

$D = 0.52$ m $\Delta p = 615$ Pa/meter

15.9

15-6

There are several ways to estimate the pressure drop that occurs in a packed tower filled with 0.0254 m Raschig ring packing under preloading conditions. One way to do this is to use the empirical relationship developed by Leva[†]. The relation in US customary units is given by

$$\Delta p / Z = \gamma (10)^{\phi L/\rho_L} G^2 / \rho_V$$

where Δp is the pressure drop in $\text{lb}_f / \text{ft}^2$, Z packed height in ft, γ, and ϕ constants dependent on the packing size and type of packing, L the superficial liquid mass velocity based on the cross-sectional area of the empty tower in $\text{lb/h}\cdot\text{ft}^2$, and ρ_L and ρ_V the density of the liquid and vapor respectively. For 0.0254m Raschig rings, $\gamma = 32\times10^{-8}$ and $\phi = 4.3\times10^{-3}$.

Since the relationship is in US customary units, it will be easier to provide the solution in those units and convert the resulting answer to SI units. Accordingly,

$$L = (1820)(2.204)/0.785D^2 = 5110/D^2 \text{ lb/h}\cdot\text{ft}^2$$

$$G = (1365)(2.204)/0.785D^2 = 3832/D^2 \text{ lb/h}\cdot\text{ft}^2$$

$$\rho_L = 1000/16.018 = 62.4 \text{ lb/ft}^3$$

$$\rho_V = (29/359)(4.92/530) = 0.075 \text{ lb/ft}^3$$

$$\Delta p / Z = 32\times10^{-8}[10^{(4.3\times10^{-3})(5110/D^2)/62.4}][(3832)/D]^2/0.075]$$

$$= \left(\frac{62.3}{D^4}\right)(10)^{0.35/D^2}$$

Product cost to overcome the Δp per foot of packed height in \$/y

$$\text{Blower cost} = \left(\frac{\Delta p}{Z}\right)\left[\frac{(1365)(2.204)(8000)}{(0.075)(2.655\times10^6)}\right](0.08) = 9.667\frac{\Delta p}{Z}$$

Set $Z = 1$ ft

$$= (9.667)(62.3/D^4)(10)^{0.35/D^2} = (605/D^4)(10)^{0.35/D^2} (\$/\text{y})/\text{ft}$$

[†] M. Leva, *Chem. Eng. Prog. Symp. Ser.*, **50**(10, 51(1954).

15-6 (continued)

Fixed charges in $(\$/\text{y})/\text{ft}$

$$\text{Packed volume} = Z(0.785D^2)$$

$$\text{Purchased tower cost} = (1413)(0.02831)(Z)(0.785D^2)$$

$$\text{Purchased packing cost} = (777)(0.02831)(Z)(0.785D^2)$$

When $Z = 1\,ft$

$$\text{Fixed charges} = (2)[(1413 + 777)(0.02831)(0.785D^2)(0.2)$$
$$= 1947D^2(\$/\text{y})/\text{ft}$$

$$\text{Total annual cost/ft} = 19.47D^2 + (605/D^4)(10)^{0.35/D^2}$$

$$\text{At optimum } D, \frac{d(\text{total annual cost}/\text{ft})}{dD} = 0$$

$$\frac{d(19.47D^2)}{dD} + \frac{d[(605/D^4)(10)^{0.35/D^2}]}{dD} = 0$$

Solving this equation shows that $D_{opt} = 2.05$ ft = **0.625 m**

Answer

This can be verified by using Mathcad as shown below:

Diameter, ft	Diameter, m	Annual cost ($ / y) / m
1	0.305	4506
1.5	0.457	705
2.0	0.610	406.80
2.02	0.616	405.80
2.05	0.625	404.50 min
2.06	0.628	407.50
2.1	0.640	408.50
2.5	0.762	457
3.0	0.914	602

15:11

15-7

Based on the concept of lowering the activity coefficient of the solute by the addition of a solvent, a table of interactions for solvent screening has been developed by Cusack, et. al.[†] which incorporate the earlier suggestions by Robbins.[‡] With the aid of this table, the following solvents are suggested (the underlined component is the solute),

Components	Solvent(s)
a. water and acetic acid	aromatic alcohols
b. water and acetone	trichloroethane, methyl isobutyl ketone, any multihalo-paraffin with an active H
c. water and aniline	methylcyclohexane
d. water and ethyl alcohol	phenol

Answer

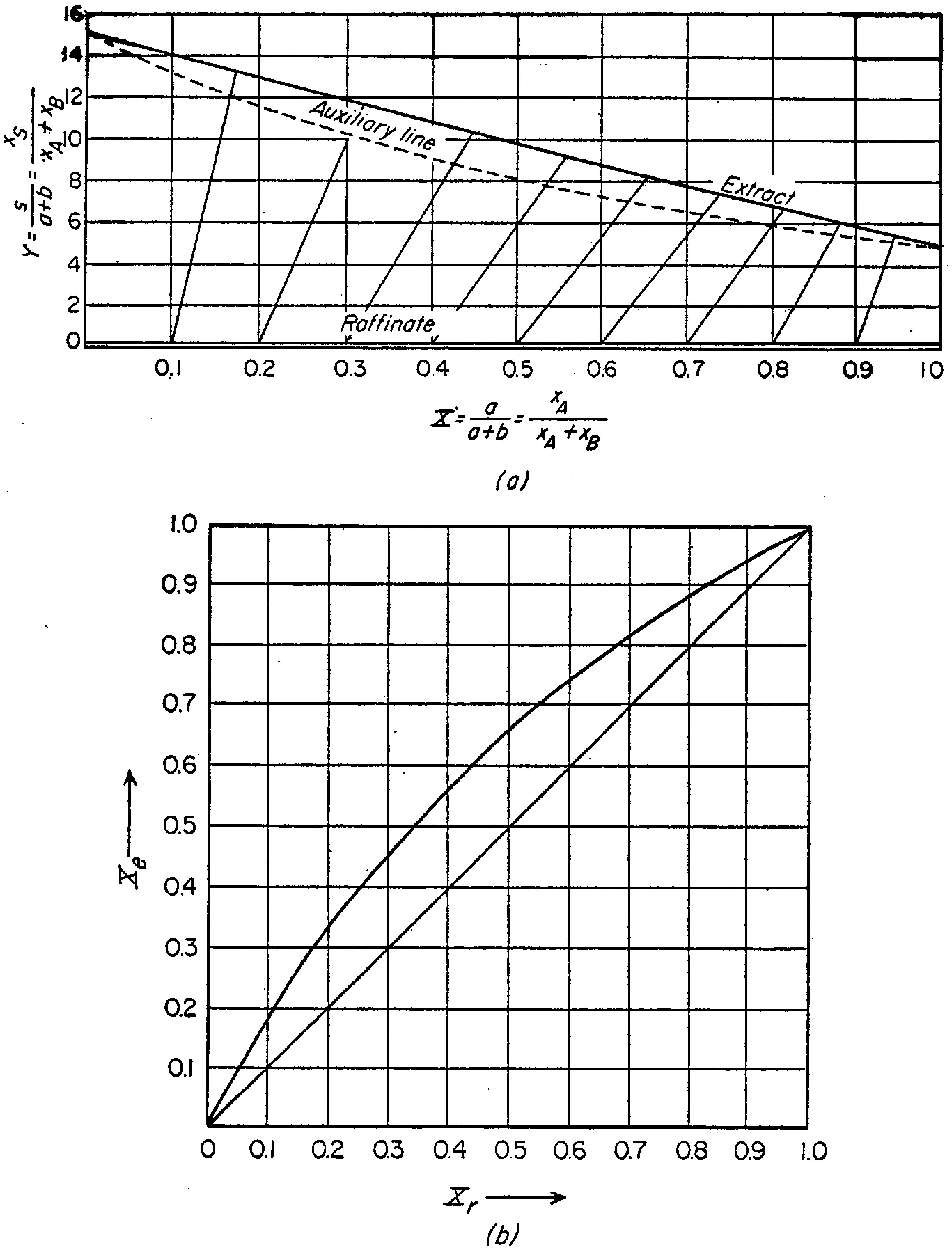
[†] R.W. Cusak, P. Fromeaux, and P. Glate, *Chem. Eng.* **98**(2), 66(1991).

[‡] L.A. Robbins, *Chem. Eng. Progr.* **76**(10), 58(1980).

15-8

Once the extraction data is obtained from the literature, this problem can be solved using either an available computer simulation software programs or developing a modified Ponchon-Savarit diagram. The latter will be used in this solution.

The extraction data at 25°C obtained by Varteressian and Fenske† are plotted below.



System aniline–*n*-heptane–MCH at 25°C: (*a*) *XY* diagram. *a*, solute, MCH. *b*, diluent, *n*-heptane. *s*, solvent, aniline. (*b*) Equilibrium, raffinate and extract, on solvent-free basis, X_e vs. X_r.

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† K.A. Varteressian and M.R. Fenske, *Ind. Eng. Chem.*, 61(7), 184 (1954).

15-8 (continued)

In the first figure X is defined as the ratio of the mass of the solute to the summed mass of the solute and the diluent. Likewise Y is defined as the ratio of mass of the solvent to the summed mass of the solute and the diluent. The second figure provides the equilibrium between the raffinate X_r and extract X_e, on a solvent-free basis.

The extract-raffinate data from the first figure is used to develop the Ponchon-Savarit diagram. The solvent content of the raffinate is low and Y_r is considered zero. X_F, X_o, and X_B are located at values of 0.40, 0.98, and 0.01, respectively. The upper operating point P_D is located using the given reflux ratio of 7.7. Thus,

$$\frac{P_D - 5}{5} = 7.7 \quad P_D = 43.5$$

The line P_DFP_B drawn through P_D and F establishes P_B to satisfy the material balance around the extraction unit. Construction of the Ponchon-Savarit diagram can now be completed with the use of equilibrium tie lines obtained from the second figure on the previous page. The fully constructed diagram is shown below showing that **26 ideal stages** are required to effect the desired separation. Feed is located on the 11th stage. **Answer**

Per 100 kg of feed, the solvent-free raffinate product is

$$[(0.98-0.40)/(0.98-0.01)]100 = 59.8\,\text{kg}$$

The solvent-free extract product is 100 – 59.8 = 40.2 kg

The solvent required, per kg of product is given by

$$(P_B - Y_r)(59.8) = (29.2-0)(59.8) = \mathbf{1745\,kg}$$

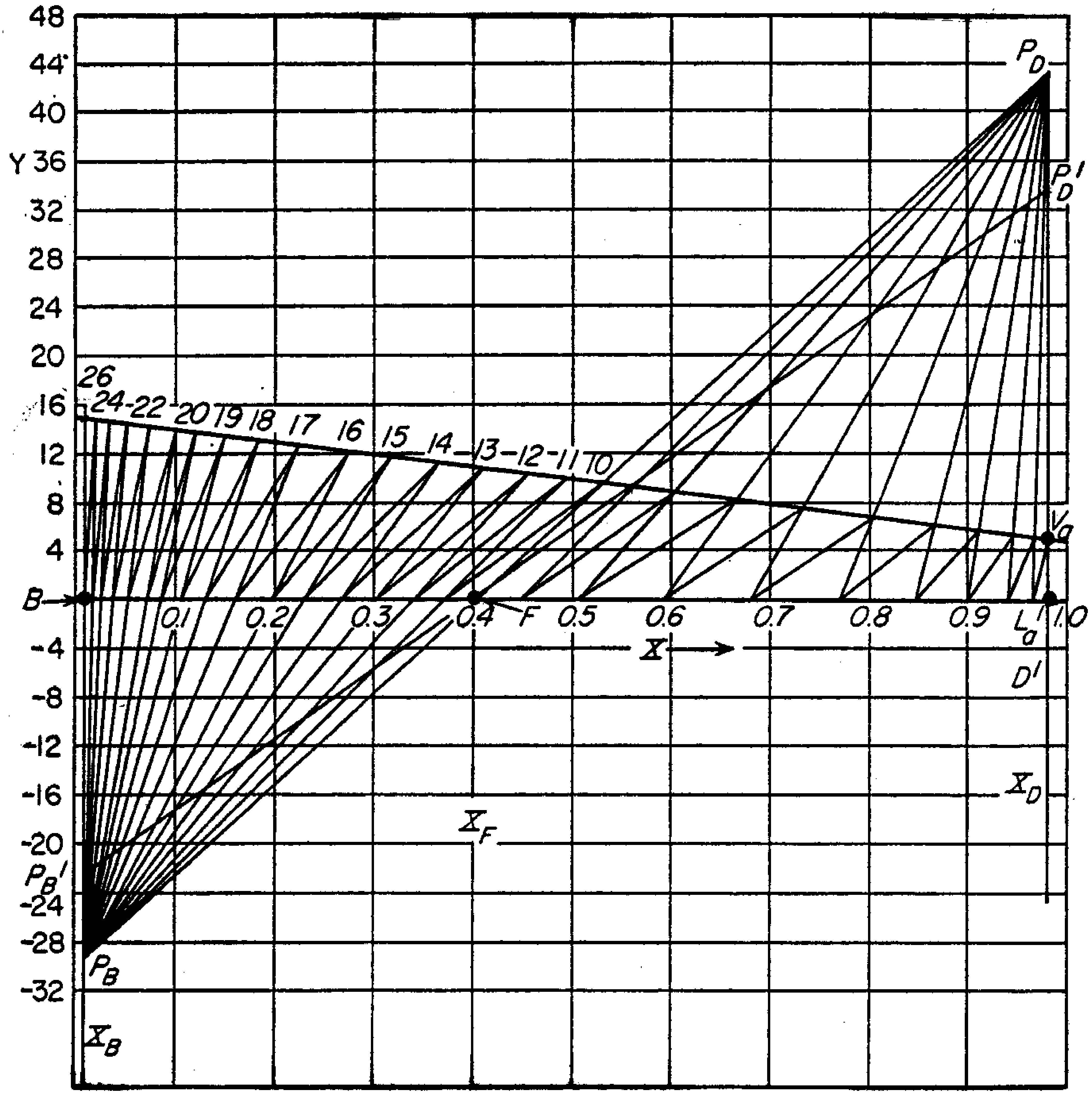
Answer

The solvent removed in the solvent separator is given by

$$(P_D - Y_r)(40.2) = (43.5-0)(40.2) = \mathbf{1745\,kg}$$

Answer

15-8 (continued)



Ponchon-Savarit diagram

15-9

From Example 15-6, $\alpha^{*}_{O_2,N_2} = 2.99$, $r = p_R / p_P = 0.1$

$$n_F = 1513\,\text{kg mol/h},\ x_{F_1O_2} = 0.21$$

To obtain α_{O_2,N_2} use Eq. (15-60) after it has been simplified to Eq. (3) of Example 15-6, namely

$$\alpha_{O_2,N_2} = \alpha^{*}_{O_2,N_2}\left[\frac{x_{F_1O_2}(\alpha_{O_2,N_2} - 1) + 1 - r\alpha_{O_2,N_2}}{x_{F_1O_2}(\alpha_{O_2,N_2} - 1) + 1 - r}\right]$$

Setting, $x_{F_1O_2} = 1$, this relation can be solved with Mathcad to obtain a value of 2.603 for α_{O_2,N_2}.

An overall module material balance on O_2 results in

$$x_{F_1O_2} n_F = x_{R_1O_2}(1 - \Theta) n_F + y_{P_1O_2} \Theta n_F$$

where $y_{P_1O_2}$ is defined in Eq. (15-66) as

$$y_{P_1O_2} = x_{R_1O_2}{}^{1/(1-\alpha)}\left(\frac{1-\Theta}{\Theta}\right)\left[\left(1 - x_{R_1O_2}\right)^{\alpha/(x-1)}\left(\frac{x_{F_1O_2}}{1 - x_{F_1O_2}}\right)^{\alpha/(\alpha-1)} - x_{R_1O_2}^{1/(1-\alpha)}\right]$$

Here α is α_{O_2,N_2} obtained above as 2.99.

For a stage cut of 0.4, the last two equations can be solved with Mathcad to give

$$x_{R_1O_2} = 0.122 \qquad y_{P_1O_2} = 0.342$$

Use Eq. (15-59a) to obtain the required area

$$A_M = \frac{y_{P_1O_2} n_P}{P_{M_1O_2}(x_{R_1O_2}\underline{p_R} - y_{P_1O_2}\underline{p_P})}$$

$$= \frac{(0.342)(1513)(0.4)}{3.50\times10^{-8}\left[0.122(1.013\times10^{6}) - 0.342(1.013\times10^{3})\right]}$$

$$= \underline{\mathbf{66{,}500\ m^2}}$$

Answer

15-10

From the information given in the problem the interparticle void fraction ε_P, the extraparticle void fraction ε, and the pore radius r_p are obtained from

$$\varepsilon_P = \left(\frac{1}{777}-\frac{1}{2178}\right)^{-1} \Big/ \left(\frac{1}{777}\right)^{-1} = 0.643$$

$$\varepsilon = \left(\frac{1}{509}-\frac{1}{777}\right)^{-1} \Big/ \left(\frac{1}{509}\right)^{-1} = 0.345$$

$$r_p = 2\varepsilon_p / A\rho_p = (2)(0.643)/(1.1\times10^{9})(777) = 1.504\times10^{-9}\,\mathrm{m}$$

The fluid phase diffusivity is $D_i = 7.42\times10^{-2}\,\mathrm{cm^2/s}$

The pore diffusivity is obtained from (Satterfield, *Mass Transfer in Heterogeneous Catalysis*, p. 43, MIT, Cambridge, MA 1970)

$$D_p = \frac{1}{\tau}\left[\left(\frac{3}{4r_p}\right)\left(\frac{\pi M}{2RT}\right)^{1/2} + \frac{1}{D_i}\right]^{-1}$$

For a tortuosity factor of 4, the D_p is $1.45\times10^{-3}\,\mathrm{cm^2/s}$

To determine whether pore diffusion is controlling, evaluate the Reynolds and Schmidt dimensionless numbers and use the figure below which relates the effect of the Re Sc product, the distribution ratio, and the diffusivity ratio on the height of a transfer unit.

$$\mathrm{Re} = Vd_p/\upsilon = (0.3)(0.0017)/(1.08\times10^{-5}) = 47.2$$

$$\mathrm{Sc} = \upsilon/D_i = (1.08\times10^{-5})/(7.42\times10^{-6}) = 1.455$$

$$\mathrm{Re\,Sc} = (47.2)(1.455) = 68.7$$

The isotherm parameters based on the feed concentration are

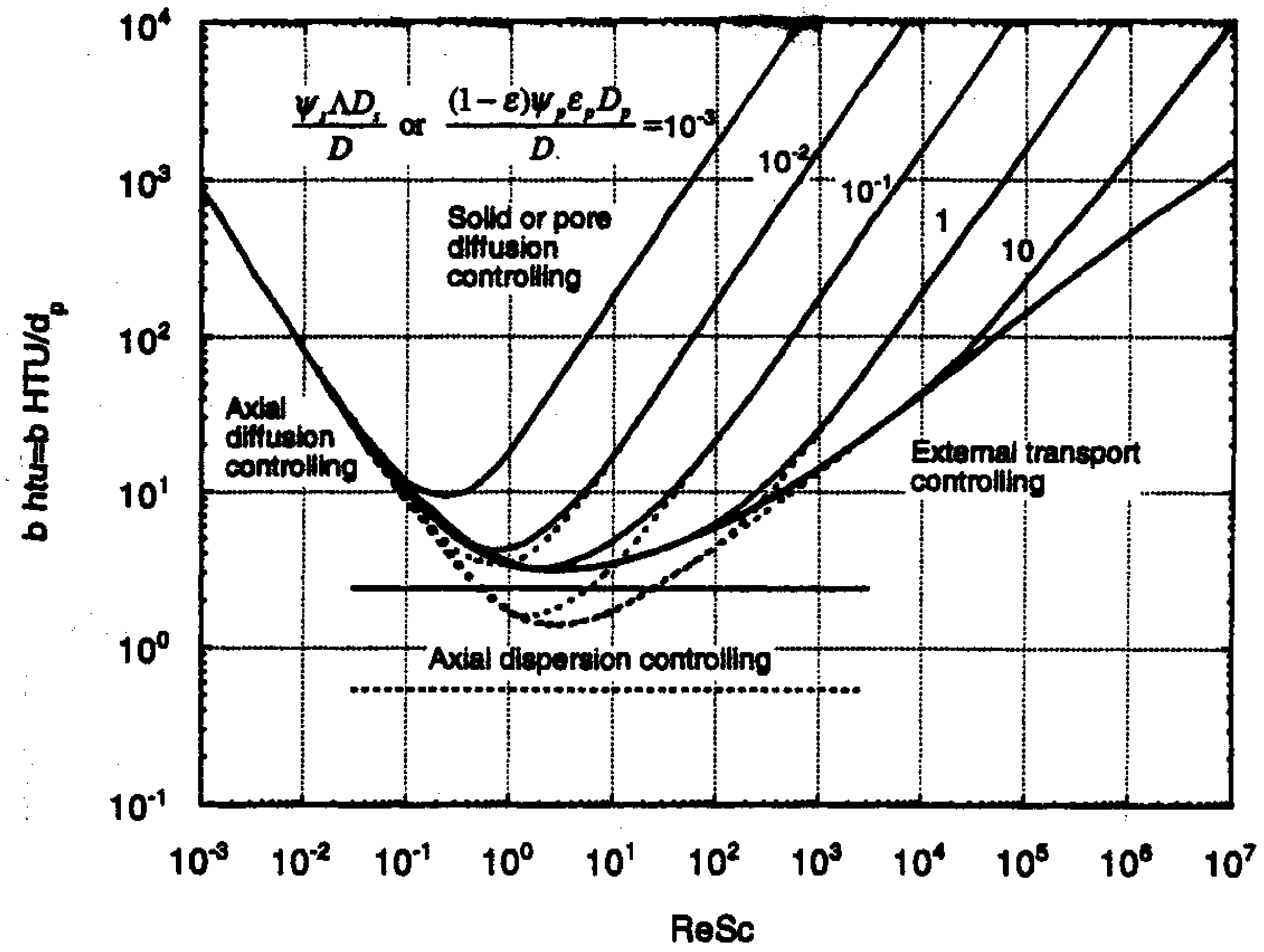
$$R = 1/(1+K_a p_a) = 1/(1+0.346(10)(0.1)) = 0.7424$$

where a correction factor Ψ_p for constant R is given by

$$\Psi_p = 0.775/(1-0.225R^{0.5}) = 0.961$$

and $(1-\varepsilon)\Psi_p\varepsilon_p D_p/D_i) = (1-0.345)(0.961)(0.643)(1.45\times10^{-3})/0.0742 = 7.93\times10^{-3}$

15-10 (continued)



Effect of *ReSc* group, distribution ratio, and diffusivity ratio on height of a transfer unit. Dotted lines for gas and solid lines for liquid-phase systems.

Location of this point at an abscissa value of 68.7 for the ReSc product, indicates that **pore diffusion** is controlling.

Answer

When pore diffusion is controlling, the expression for the rate coefficient can be approximated from *Perry's Chemical Engineers' Handbook*, 7th ed., Table 16-2 as

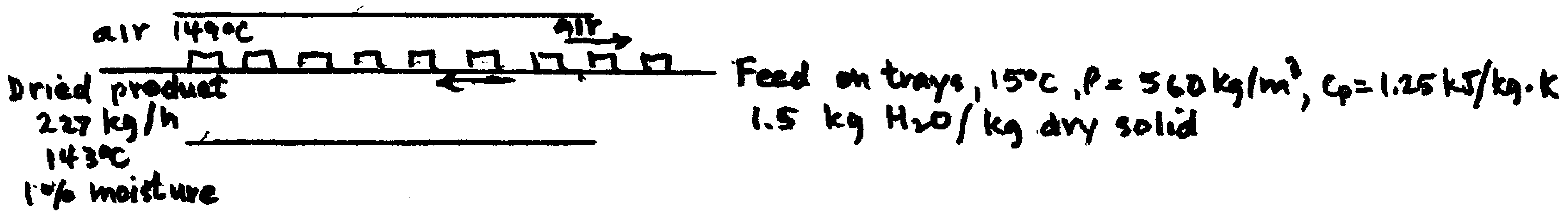
$$k = 15\Psi_p(1-\varepsilon)\varepsilon_p D_p / \Lambda r_p^2$$

Substituting appropriate values in this relation with the isotherm parameter $\Lambda = 9.56$ results in

$k = 0.13\ s^{-1}$

Answer

15-11



In this drying process assume shrinkage during drying is negligible. Also assume operation at 1 atm.

Heat transfer requirement is approximated from the following heat balance for the water and the solid:

Heat to evaporate moisture (C_{p,H_2O} vapor $= 1.884\,\mathrm{kJ/kg\cdot K}$)

$$\dot{q}_{H_2O} = 227(1.5-0.01)[(100-15)(4.186)+2256+1.884(60-100)]$$
$$= 887{,}887 \text{ or } 857{,}900\,\mathrm{kJ/h}$$

Heat of solid

$$\dot{q}_{\text{solid}} = (1.25)(227)(143-15) = 36{,}200\,\mathrm{kJ/h}$$
$$\dot{q}r = 857{,}900+36{,}200-894{,}100\,\mathrm{kJ/h}$$

Relative humidity of entering air with 0.03 kg H_2O/kg dry air at 149°C is ~1%

Required mass of air is

$$\dot{m}_{air} = 894{,}100/(1.06)(149-60) \qquad \text{where } 1.06\ \mathrm{kJ/kg\cdot K}$$
$$= 9478\ \mathrm{kg/h} \qquad \text{is the humid heat at 149°C and 1\% RH}$$

Rate of water removal

$$\dot{m}_{H_2O} = (227)(1.5-0.01)-338.2\,\mathrm{kg/h}$$

Humidity of air leaving the dryer $= 0.03 + 338.2 / 9478$

$= 0.0656$ kg water/kg dry air

15-11 (continued)

This exit humidity corresponds to a relative humidity of 48% and 60°C. The wet bulb temperature at this condition is

$$T_{wb} = 46.7^\circ\text{C}$$

Mass of air recirculated

$$\dot{m}_{air} = 9478(0.03 - 0.01)/(0.0656 - 0.01)$$
$$= 3410\,\text{kg/h}$$

The heat required for the constant rate section is 642,000 kJ/h

Air temperature where surface drying stops is

$$T = 60 + (149 - 60)(642,000/894,100) = 123.9^\circ C$$

The number of transfer units (NTU) is defined by

$$NTU = \ln \frac{T_1 - T_{wb}}{T_c - T_{wb}}$$

where T_1 = air temperature entering the constant-rate zone

T_c = air temperature at the critical moisture content

$$NTU = \ln \frac{123.9 - 46.7}{60 - 46.7} - \underline{\mathbf{1.75}}$$

Answer

Under normal tray drying conditions, NTU in the falling rate section is obtained from

$$NTU = \frac{T_c - T_2}{\Delta T_m} \qquad \text{where} \qquad \Delta T_m = \frac{(123.9 - 46.7) - (149 - 143)}{\ln(123.9 - 46.7)/(149 - 143)}$$

$$= \frac{(149 - 123.9)}{27.9} \qquad\qquad = 27.9^\circ\text{C}$$

$$= \underline{\mathbf{0.90}}$$

Answer

$NTU_{total} = 1.75 + 0.90 = 2.65$

15-11 (continued)

The open cross-sectional area of the tunnel drier is given by

$$A = 9478/(2.71)(3600) = 0.97\,\mathrm{m}^2$$

The length of the dryer is obtained with the empirical equation

$$L = (72.77)(0.0381)(2.71)^{0.2}(2.65) = 8.96\,\mathrm{m} \text{ or } \underline{\mathbf{9m}}$$

Answer

15'.21

15-12

In this evaporation process assume a temperature driving force of 5°C between the wall temperature and the evaporating water, i.e., the T_W = 105°C. Since the Reynolds number of the liquid flow down the vertical will change, it will be necessary to evaluate values at both ends to arrive at an average heat transfer coefficient.

The mass flow of water per tube periphery is given by

$$\Gamma_o = \dot{m}/\pi D_i = 0.01/\pi(0.05) = 0.0637\,\text{kg/m}\cdot\text{s}$$

The Reynolds number is given by $4\Gamma_o/\mu_L$. At the periphery

$$\text{Re}_{in} = (4)(0.0637)/2.79\times10^{-4} = 913.3 \quad \mu_L = 2.79\times10^{-4}\,\text{Pa}\cdot\text{s}$$

The film is laminar, but waves are on the surface. If laminar flow in this wavy regime (30<Re<1600) continues down to the bottom of the tube, Kutateladze (*Fundamentals of Heat Transfer*, Academic Press, New York, 1963) developed the following relation to determine the Reynolds as a function of distance down the tube:

$$\text{Re}_{\text{out}}^{1.22} = \text{Re}_{\text{in}}^{1.22} - 3.69\frac{k_L(T_W - T_S)L}{\Delta H_V \mu_L}\left[\frac{\rho_L(\rho_L - \rho_V)g}{\mu_L^2}\right]^{1/3}$$

Properties required are: $k_L = 0.68$ W/m·K; $\Delta H_V = 2.03\times10^6$ J/kg

$\rho_L = 958\,\text{kg/m}^3$; $\rho_V = 0.6\,\text{kg/m}^3$

Substituting these values and using Mathcad to solve for Re at the lower end of the tube

$$\text{Re}_{\text{out}}^{1.22} = (913.3)^{1.22} - 3.69\frac{(0.68)(105-100)3}{(2.03\times10^6)(2.79\times10^{-4})}\left[\frac{958(958-0.6)9.8}{(2.79\times10^{-4})^2}\right]^{1/3} = 856$$

$\text{Re}_{\text{out}} = 253.3$; this indicates continued wavy laminar flow

The fraction of water evaporated is ($\text{Re}_{\text{min}} - \text{Re}_{\text{out}}$) / Re_{in} = 0.726. Since the average heat transfer coefficient can be related to the inlet and exit Reynolds numbers as

$$\bar{h} = \frac{(\text{Re}_{\text{in}} - \text{Re}_{\text{out}})}{-\int_{\text{Re}_{\text{in}}}^{\text{Re}_{\text{out}}} d\,\text{Re}/h}$$

15-12 (continued)

An expression can be formulated for the average heat transfer assuming that T_W and T_S remain constant down the length of the tube. Kutateladze has developed such a relation

$$\bar{h} = 0.922 k_L \left[\frac{\rho_L(\rho_L - \rho_V)g}{\mu_L^2}\right]^{1/3} \frac{(\text{Re}_{\text{in}} - \text{Re}_{\text{out}})}{(\text{Re}_{\text{in}}^{1.22} - \text{Re}_{\text{out}}^{1.22})}$$

$$= (0.922)(0.68)\left[\frac{(958)(958-0.6)(9.8)}{(2.79\times10^{-4})}\right]^{1/3}\left[\frac{913.3-253.3}{(913.3)^{1.22}-(253.3)^{1.22}}\right]$$

$$= 6.23\times10^3\ \text{W/m}^2\cdot\text{K}$$

The volumetric rate of steam can now be evaluated. The total rate of heat transfer is

$$\dot{q} = \bar{h}(T_W - T_S)\pi D L$$

$$= (6.23\times10^3)(105-100)\pi(0.05)(3) = 1.47\times10^4\text{W}$$

The volume of steam generated per second is

$$v_{\text{steam}} = \dot{q}/\Delta H_V \rho_V = 1.47\times10^4/(2.03\times10^6)(0.6)$$

$$= \underline{\mathbf{0.012\ m^3/s}}$$

Answer

The maximum velocity of the steam is at the top of the tube. Thus,

$$V_{\max} = 0.012/(0.785)(0.05)^2 = 6.1\ \text{m/s}$$

This is about 22% of the flooding velocity of 27 m/s, but even this velocity can have some effect on the interfacial shear stress. **Answer**

15-13

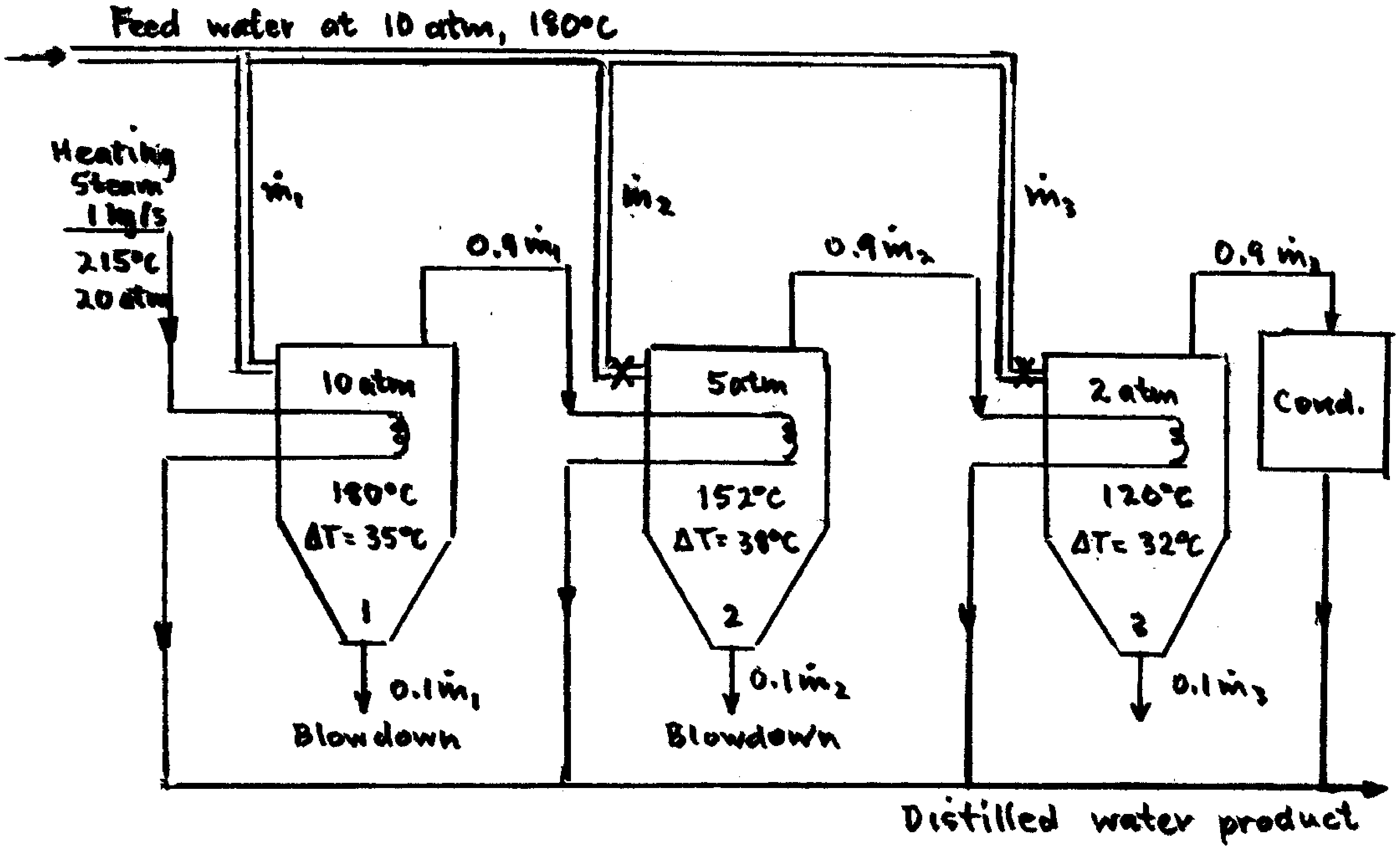


Diagram of triple-effect evaporator

To obtain essentially equal temperature-differences over the triple effect evaporator, operate the first evaporator at 10 atm, the second at 5 atm, and the third at 2 atm. (The latter choice was made to maintain a positive driving force from the exit from the evaporator.) The driving temperature differences are

Stage 1: $\Delta T_1 = 215 - 180 = 35^\circ\text{C}$

Stage 2: $\Delta T_2 = 180 - 152 = 28^\circ\text{C}$

Stage 3: $\Delta T_3 = 152 - 120 = 32^\circ\text{C}$

15-13 (continued)

Values of enthalpy for steam and water for the 3 stages used in the solution

	Steam	Stage 1	Stage 2	Stage 3
Pressure, atm	20	10	5	2
Temp., °C	215	180	152	120
h_V, kJ/kg $\times 10^{-3}$		2.778	2.749	2.706
h_L, kJ/kg $\times 10^{-3}$		0.763	0.641	0.504
ΔH_V, kJ/kg $\times 10^{-3}$	1.881	2.015	2.108	2.202

For stage 1, assume a steam feed of 1 kg/s is condensed at 20 atm and that the liquid being blown down has the same enthalpy $h_{1,L}$ as that entering. The heat balance for stage 1 is

$$\dot{q}_1 + \dot{m}_1 h_{1,L} = 0.9\dot{m}_1 h_{1,V} + 0.1\dot{m}_1 h_{1,L}$$

$$\dot{q}_1 = 0.9\dot{m}_1 \Delta H_{1,V}$$

$$\dot{m}_1 = \dot{q}_1 / 0.9\Delta H_{1,V} = 1.881\times10^3/(0.9)(2.015\times10^3) = 1.037\,\text{kg/s}$$

For stage 2, $\dot{m}_{\text{in}} = (0.9)(1.037) = 0.9345\,\text{kg/s}$ and is a source of heat for evaporation. The 1 kg/s of condensed steam from stage 1 can be added to the distilled water product. Assume in the heat balance for stage 2, all steam coming from stage 1 to stage 2 is condensed. Thus,

$$\dot{q}_2 = (0.9335)(2.015\times10^3) = 1.881\times10^3\,\text{kJ/s}$$

which is identical to $\dot{q}_1$ since the liquid enters and leaves stage 1 with the same enthalpy.

Since the pressure in stage 2 is 5 atm, some of the feedwater entering is vaporized because of the isenthalpic pressure reduction. The overall energy balance for stage 2 is

$$\dot{q}_1 + \dot{m}_2 h_{1,L} = 0.9\dot{m}_2 h_{2,V} + 0.1\dot{m}_2 h_{2,L}$$

$$\dot{m}_2 = \dot{q}_2 / (0.9h_{2,V} + 0.1h_{2,L} - h_{1,L})$$

$$= 1.881\times10^3 / [(0.9)(2.749) + (0.1)(0.641) - (0.763)]10^3$$

$$= 1.06\ \text{kg/s}$$

15-13 (continued)

For stage 3, $\dot{m}_{\text{in}} = (0.9)(1.06) = 0.954\,\text{kg/s}$ and the rate of condensate to the product water line is 0.9345 kg/s. The energy balance for this stage is

$$\dot{q}_3 + \dot{m}_3 h_{1,L} = 0.9\dot{m}_3 h_{3,V} + 0.1\dot{m}_3 h_{3,L}$$

$$\dot{m}_3 = \dot{q}_3 / 0.9h_{3,V} + 0.1h_{3,L} - h_{1,L}$$

$$\dot{m}_3 = (0.954)(2.202\times10^6)/[(0.9)(2.706)+(0.1)(0.504)-0.763]10^3$$

$$= 1.220 \text{ kg/s}$$

Steam is produced at a rate of (0.9)(1.220) or 1.098 kg/s and this is condensed in the condenser before being added to the distilled water product.

The total amount of distilled water produced is the sum of the water obtained from the three stages plus the water from the condenser. This total is

Stage 1	1.000 kg/s
Stage 2	0.934
Stage 3	0.954
Condenser	1.098
	3.986 kg/s
	Answer

If the feedwater at 10 atm had been flashed down to the final product pressure of 2 atm in the third stage, the fraction of steam generated would have been

$$x = (h_{1,L} - h_{3,L}) / \Delta H_{3,V}$$

$$= (0.736 - 0.504)10^3 / 2.202\times10^3$$

$$= 0.105$$

The amount of additional distilled water product would be

$$\dot{m} = (0.105((3.317) = \underline{\mathbf{0.349\ kg/s}}$$

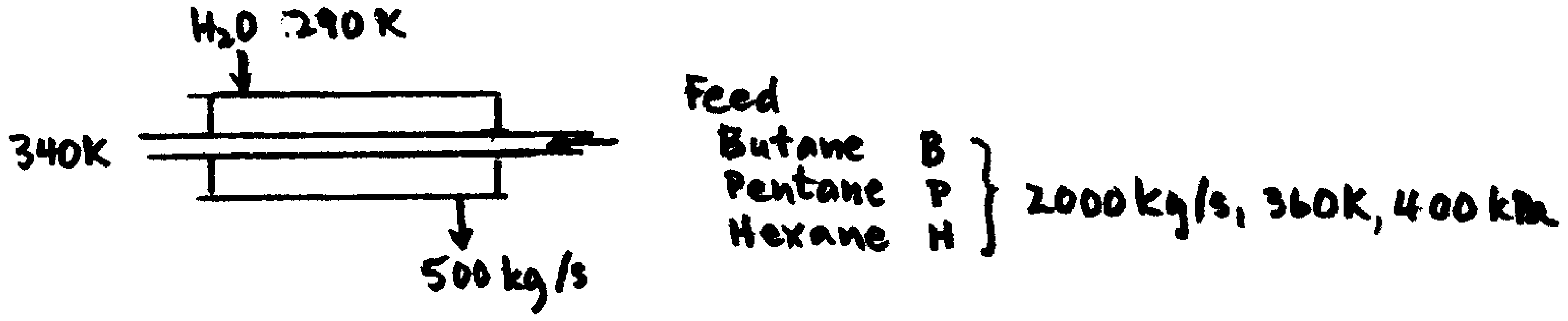
Answer

15-13 (continued)

where 3.317 kg/s is the sum of the entering feedwater along with the 1 kg/s of entering steam. Blowdown rate is 10% of the feedwater rate.

15:27

15-14



To determine the heat removed in the condensation process requires establishing the vapor-liquid relationship across the heat exchanger. The rate of heat removal changes because of a variation in the heat transfer coefficient. To approximate this change will require dividing the heat exchanger into several smaller sections. One convenient way of doing this is to use 5°C temperature intervals to divide the exchanger into four sections. Establishing the amounts of vapor and liquid of each component at these temperature intervals requires the use of equilibrium constants. The solution at 360K will be shown below.

Convert feed to mols (B refers to butane, P to pentane, H to hexane)

$\dot{n}_B = (200/3)(58.1) = 1.147$ kg mol/h

$\dot{n}_P = (200/3)(72.1) = 0.924$ kg mol/h

$\dot{n}_H = (200/3)(86.2) = \underline{0.773 \text{ kg}}$ mol/h

Total 2.844

From equilibrium relations in mass units, the following expression relates the ratios of mols of gas to mols of liquid at a specific temperature.

$$\dot{n}_T = \left(1+\frac{\dot{n}_V}{\dot{n}_L}\right)\sum_{i=1}^{n}\frac{\dot{n}_T}{K(\dot{n}_V/\dot{n}_L)+1} \quad \text{and } \dot{n}_T = \dot{n}_V + \dot{n}_L$$

at 360 K

$K_B = 3.108$ $K_P = 1.105$ $K_H = 0.458$

$$2.844 = \left(1+\frac{\dot{n}_V}{\dot{n}_L}\right)\left[\frac{2.844}{3.108(\dot{n}_V/\dot{n}_L)+1}+\frac{2.844}{1.105(\dot{n}_V/\dot{n}_L)+1}+\frac{2.844}{0.458(\dot{n}_V/\dot{n}_L)+1}\right]$$

15-14 (continued)

Use of Mathcad quickly determines $\dot{n}_V$ and $\dot{n}_L$ for each component.

Results from the calculation for a temperature of 360K

	$\dot{n}_L$	$\dot{n}_V$	$\dot{n}_T$
Butane, kg mol/h	0.0117	1.1353	1.147
Pentane, kg mol/h	0.0261	0.8979	0.924
Hexane, kg mol/h	0.0507	0.7223	0.773
	0.0885	2.7555	2.844

To obtain the energy content of this mixture at 360K, requires use of molar enthalpies relative to a reference temperature (340K in this problem) obtained from Maddox, *Heat Exchanger Design Handbook*. Units are in kJ/kgmol.

	Butane		Pentane		Hexane	
Temp.	Liq.	Vap.	Liq.	Vap.	Liq.	Vap.
340	0	18,075	0	23,940	0	28,820
345	813	18,336	945	24,502	1,031	29,568
350	1,569	18,773	1,976	25,057	2,180	30,318
355	2,441	19,121	2,734	25,620	3,240	31,068
360	3,371	19,586	3,579	26,176	4,167	31,818

The energy content in kJ/kg at 360K is then

$$h_{\text{mean}} = \frac{1}{\dot{m}_T}\sum_{i=1}^{n}(h_B\dot{n}_B + h_P\dot{n}_P + h_H\dot{n}_H)_V + (h_B\dot{n}_B + h_P\dot{n}_P + h_H\dot{n}_H)_L$$

$$= \frac{1}{200}[19,586)(1.1353) + (26,176)(0.8979) + (31,818)(0.7223) +$$

$$(3,371)(0.0117) + (3,579)(0.0261) + (4,167)(0.0507)]$$

$$= 345.4\,\text{kJ/kg}$$

Repeating this same procedure at a temperature of 340K provides a h_{mean} of 27.5 kJ/kg. Thus, the heat removed in the condenser is

$$\dot{q}_r = 200(345.4 - 27.5) = 63,580\,\text{kJ/s}$$

15-14 (continued)

The calculations of 340K, 345K, 350K, 355K and 360K can be obtained by use of a spreadsheet. The results are tabulated below:

	340K	345K	350K	355K	360K
$\dot{n}_{B,L}$ kg mol/s	0.9532	0.5572	0.3070	0.1269	0.0117
$\dot{n}_{P,L}$ kg mol/s	0.8631	0.6739	0.4701	0.2373	0.0261
$\dot{n}_{H,L}$ kg mol/s	0.7549	0.6778	0.5562	0.3452	0.0507
$\dot{n}_{B,V}$ kg mol/s	0.1939	0.5898	0.8392	1.0201	1.135
$\dot{n}_{P,V}$ kg mol/s	0.0609	0.2500	0.4539	0.6867	0.898
$\dot{n}_{H,V}$ kg mol/s	0.0187	0.0957	0.2173	0.4193	0.722
h_V, kJ/kg	27.5	108.1	181.7	261.1	345.4
$\dot{q}$, kJ/s	-63,580	-47,450	-32,740	-16,850	0

The outlet temperature of the cooling water can now be determined.

$$\begin{aligned} T_{H_2O,\text{out}} &= T_{H_2O,\text{in}} + \dot{q}_T / \dot{m}_{H_2O} C_{pH_2O} \\ &= 290 + 63{,}850/(500)(4.2) = 290 + 30.3 \\ &= 320.3\,\text{K} \end{aligned}$$

The inlet vapor mass flow rate is given by

$$\dot{m}_V = (1.135)(58.1) + (0.989)(72.1) + (0.722)(86.2) = 193\,\text{kg/s}$$

Similarly the outlet vapor mass flow rate using information from the table is determined as 17.3 kg/s.

Since the vapor flows within the tube, assume that the vapor heat transfer coefficient varies with $\text{Re}^{0.8}$ or $\dot{m}_V^{0.8}$. Thus, at the outlet,

$$\begin{aligned} h_{V,\text{out}} &= h_{V,\text{in}} (\dot{m}_{V,\text{out}} / \dot{m}_{V,\text{in}})^{0.8} = (150)(17.3/193)^{0.8} \\ &= 21.8\,\text{W/m}^2 \cdot \text{K} \end{aligned}$$

This same calculation can be repeated for each temperature increment along the condensation curve occurring within the heat exchanger. The results of these calculations are tabulated below:

15-14 (continued)

Temp, K (mixture)	Temp, K (water)	$\dot{m}_V$, kg/s	h_V, W/m$^2\cdot$K
340	290	17.3	21.8
345	297.7	60.6	59.3
350	304.7	100.2	89
355	312.3	144.9	119
360	320.3	193	150

Additional corrections to h_V can be made because of condensation conditions, but will not be included in this solution. With the establishment of the four temperature sections in the heat exchanger, the Silver-Bell-Ghaly method can now be used to approximate the required areas in these four sections. Consider the increment from 360K to 355K. The $\Delta\dot{q}$ for this section is 16,850 kJ/s. This can be approximated in this section by

$$\Delta\dot{q}_V = \dot{m}_{V,\text{avg}} C_p \Delta T$$

where $\dot{m}_{V,\text{avg}} = (193+144.9)/2 = 168.95\,\text{kg/s}$

$$\Delta\dot{q}_V = (168.95)(2)(5) = 1689.5\,\text{kJ/s}$$

Thus, $\dfrac{\Delta\dot{q}_V}{\Delta\dot{q}} = \dfrac{1689.5}{16,850} = 0.1003$

This ratio is inserted in the equation used in the method to obtain the increment of area required in this section. This relation is

$$\Delta A = \Delta\dot{q}\left(1+\frac{\Delta\dot{q}_V}{\Delta\dot{q}}U'/\bar{h}_V\right)/U'(\bar{T}_E-\bar{T}_C)$$

where $\bar{T}_E$ is the mean equilibrium temperature for the increment, $\bar{T}_C$ the mean temperature of the cooling liquid, U' the overall heat transfer coefficient from the coolant to the condensate film interface and $\bar{h}_V$ the mean vapor heat transfer coefficient. For this section

$$\bar{T}_E = (360+355)/2 = 357.5\text{K}$$

$$\bar{T}_C = (320.3+312.3)/2 = 316.3\text{K}$$

15-14 (continued)

$$U' = 800\ \text{W/m}^2\cdot\text{K}$$

$$\bar{h}_v = (150 + 119)/2 = 134.5\ \text{W/m}^2\cdot\text{K}$$

Substituting these values in the incremental area equation

$$\Delta A = 16.85\times10^6[1+(0.1003)(800)/134.5]/((800)(357.5-316.3)$$
$$= 816\,\text{m}^2$$

Similar calculations for the three other sections shows required areas of 722, 633, and 611 m^2 for a total of

$$A_T = 816 + 722 + 633 + 611 = \underline{\mathbf{2782\ m^2}}$$

Answer

Using one of the available computer simulation programs shows a requirement of 3090 m^2. However, this program includes a safety factor to recognize fouling particularly by the water, use of commercial lengths and diameters, and inclusion of construction details.

15-15

For a rotary filter, the volume of filtrate per unit time is given by Eq. (15-101b)

$$\frac{\text{Volume of filtrate}}{\text{time}} = A_D\left(\frac{2\Psi_f N_r \Delta p}{\alpha w \mu}\right)^{1/2}$$

This equation can be used for both the small and large rotary filter

Small filter: $$0.3 = 0.75\left(\frac{2(0.2)(2)(140)}{\alpha w \mu}\right)^{1/2} \qquad (1)$$

Large filter: $$3 = A_D\left(\frac{2(0.2)(1.5)(100)}{\alpha w \mu}\right)^{1/2} \qquad (2)$$

Since α, w, and μ remain constant in both filters, Eq. (2) can be divided by Eq. (1) to obtain the new area of the filter drum

$$\frac{3 = A_D\left[(2)(0.2)(1.5)(100)\right]^{1/2}}{0.3 = 0.75\left[(2)(0.2)(2)(140)\right]^{1/2}}$$

A_D = **10.25 m^2**

Answer

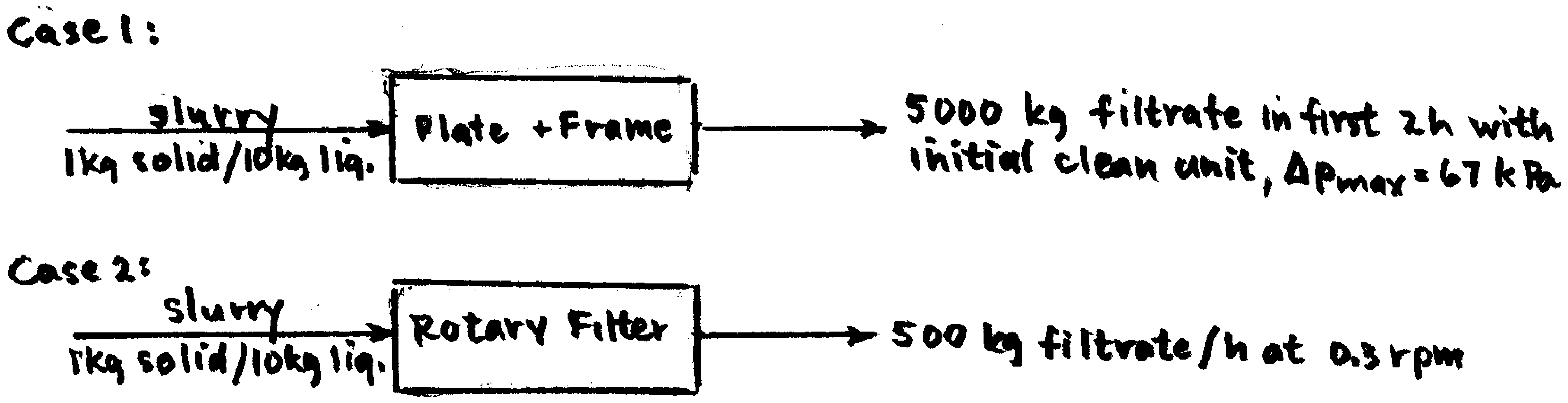
Purchased cost of filter (Jan. 2002) from Figure 15-39 is given as **$120,000**

Answer

15:33

15-16

Conditions for the two filtration systems:



In each case both Ψ and Δp are constant and provide negligible filter-unit resistance.

For case 1, let w_F = kg filtrate rate/h and $\rho_F = \text{kg/m}^3$

$$w_F / \rho_F = \upsilon, \text{m}^3/\text{h}$$

From Eq. (15-91) $\qquad\qquad \Delta p = 67\,\text{kpa},\ A = 25\text{m}^2$

$$\upsilon^2 = \frac{2\Delta p A^2}{\alpha' w \mu}\Theta = \left(\frac{w_F}{\rho_F}\right)^2$$

$$\frac{(5000)^2}{2} = 1.25\times10^7 = \frac{2\Delta p A^2 \rho_F^2}{\alpha' w \mu} \qquad \text{since } \Theta = 2\text{h}$$

$$w_F = (1.25\times10^7\Theta)^{1/2}$$

$$\text{Product as kg of filtrate/24 h} = (12.5\times10^6\Theta)^{1/2}\left(\frac{24}{\Theta+3}\right)$$

$$= \frac{8.485\times10^4\Theta^{1/2}}{\Theta+3}$$

At Θ_{opt} for maximum production per day, set

$$\frac{d(\text{product/24h})}{d\Theta} = 0 \text{ and solve for } \Theta_{\text{opt}} \text{ with a result that } \Theta_{\text{opt}} \cong 3\text{h}$$

15-16 (continued)

$$\text{Maximum product/24 h} = \frac{8.485\times10^4(3)^{1/2}}{3+3} = 2.45\times10^4\,\text{kg}/24\text{h}$$

For equal product rate in the rotary filter per hour requires

$$\dot{m} = 2.45\times10^4/24 = 1020\,\text{kg/h}$$

From Eq. (15-101b)

$$\text{Volume of filtrate/unit time} = A_D\left(\frac{2\Psi_f N_r \Delta\rho}{\alpha w \mu}\right)^{1/2}$$

$$\text{kg of filtrate/h} = A_D\rho_F\left(\frac{2\Psi_f \Delta p}{\alpha w \mu}\right)^{1/2}(N_r)^{1/2}$$

Where N_r is in revolutions /h

From the data for the operating rotary filter

$$500\,\text{kg/h} = A_D\rho_F\left(\frac{2\Psi_f \Delta p}{\alpha w \mu}\right)^{1/2}[(0.3)(60)]^{1/2} \tag{1}$$

For the required rotary filter

$$1020 = A_D\rho_L\left(\frac{2\Psi_f \Delta p}{\alpha w \mu}\right)^{1/2}(N_r)^{1/2} \tag{2}$$

Divide Eq. (2) by Eq. (1), cancel common terms and solve for N_r

$$\frac{1020}{500} = \frac{(N_r)^{1/2}}{[(0.3)(60)]^{1/2}}$$

$N_r = 74.9$ r/h or **<u>1.25 rpm</u>**

Answer

APPENDIX C

Practice Session Problems

Pr. Ses. Prob. 1

Basis: 1 hr.

Moles pure H_2 in = $\begin{array}{c|c|c} 10,000,000 & 72.5 & \\ \hline 359 & 100 & 24 \end{array}$ = 842 moles

Moles pure CH_4 in = $\begin{array}{c|c|c} 10,000,000 & 27.5 & \\ \hline 359 & 100 & 24 \end{array}$ = 319 "

Total 1161 "

X = lb act. C necessary per bed.

$$0.95 = \frac{842 - 0.00838\,(0.032)\,X}{842 - 0.00838\,(0.032)\,X + 319 - 0.00838\,(0.968)\,X}$$

X = 33,900 lb. charcoal per bed

Amt. pure H_2 prod = 842 − 0.00838 (0.032)(33,900) = 832.8 moles

Amt. pure CH_4 prod. = 319 − 0.00838 (0.968)(33,900) = 44 moles

Total 876.8

Check: $\frac{832.8}{876.8}(100) = 95\%$

Cu. ft. of 95% H_2 (at S.C.) produced per

year = $\begin{array}{c|c|c|c} 876.8 & 359 & 24 & 350 \\ \hline & & & \end{array}$ = 2,650,000,000 ft^3

Cost Accounting:

Charcoal initial cost = (33,900)(3)($3.65) = $371,200

Charcoal makeup cost = (0.15)($371,200) = $55,700

Capital investment = $4,750,000 + $371,200 = $5,121,200.

Pr. Scs. Prob. 1 (Continued)

Cost Per Year

$$\text{Amortization} = \frac{1}{5}(\$5{,}121{,}200) = \$1{,}024{,}000$$

$$\text{Other operating costs} = \$2{,}250{,}000 + \$55{,}700 = \underline{\$2{,}305{,}700}$$

$$\text{Total cost per year} = \$3{,}329{,}700$$

Cost per 1000 ft^3 (at s.c.) of 95% H_2 =

$$\frac{\$3{,}329{,}700}{2{,}650{,}000{,}000}\bigg|\, 1000 = \$1.256 \approx \underline{\underline{126\ ¢/1000\ ft^3\ (s.c.)}}$$

Answer

Pr. Ses. Prob. 2

Basis: 1 hr.

Moles pure H_2 in $= \begin{array}{c|c} 10{,}000{,}000 & 0.725 \\ \hline 359 & 24 \end{array} = 842$ moles

Moles pure CH_4 in $= \begin{array}{c|c} 10{,}000{,}000 & 0.275 \\ \hline 359 & 24 \end{array} = 319$ moles

X = lb. carbon necessary/bed Total 1161

$$0.95 = \frac{842 - (0.00632)(0.032)X}{842 - (0.00632)(0.032)X + 319 - (0.00632)(0.968)X}$$

$X = 45{,}000$ lb. charcoal/bed

Moles H_2 produced $= 842 - 0.000202X = 832.9$ moles

Moles CH_4 produced $= 319 - 0.00612X = 44$ "

Total 876.9 "

Volume of ads. bed $= \begin{array}{c|c} 45{,}000 & \\ \hline 0.30 & 62.42 \end{array} = 2400$ ft^3

Find limiting gas flow - (based on max. vel. of 1 ft/sec)

H_2 Feed gas rate $= \begin{array}{c|c|c|c} 10{,}000{,}000 & & 14.7 & 540 \\ \hline 24 & 3600 & 400 & 492 \end{array} = 4.67$ act. ft^3/sec.

Regen. + Flue gas rate $= \begin{array}{c|c|c|c|c|c} 80+30 & 45{,}000 & 14.7 & 1060 & & \\ \hline & & 19.7 & 492 & 3600 & 8 \end{array}$

$= 277$ ft^3/sec

Thus, the regeneration cycle puts the most gas through per second, and this rate will set the necessary column diameter.

At a max. gas velocity of 1 ft/sec, the cross-sectional area (total) per bed must be 277 ft^2.

Pr. Ses. Prob. 2 (Continued)

Determination of Individual No. of Towers per bed:

Dia., ft	X-Sect. area/tower ft²	No. of Towers to Give 277 ft² Area	Tower Ht. ft	\$/ft. length
6	28.2	10	8.52(+10)	1170
9	63.6	5	7.55 (+10)	2140
12	113.1	3	7.07 (+10)	3280
15	176.5	2	6.81 (+10)	5840

Dia. ft	Skirt cost \$	Cost, \$/Tower	Total cost, \$ (3 beds)
6	1950	21,670 + 1950	236,200 × 3
9	3240	37,560 + 3240	204,000 × 3
12	4540	56,020 + 4540	181,700 × 3
15	5200	98,100 + 5200	206,600 × 3

Sample Calc. $D = 6$ ft. $\text{Area} = (0.785)(6)^2 = 28.2 \text{ ft}^2$

$$\text{No. of towers} = \frac{277}{28.2} = 10 \text{ (rounded off)}$$

$$\text{Tower Ht.} = \frac{2400}{(28.2)(10)} = 8.52 \text{ ft.} \qquad \text{Cost/tower} = \$1950 + (18.52)(\$1170) = \$23,620$$

$$\text{Total cost (3 beds)} = (\$23,620)(10)(3) = \$236,200 \times 3$$

Part a: 3 Towers per bed for minimum cost
Answer

Part b: Tower diameter = 12 ft.
Tower height = 7.1 ft. Answer

Pr. Ses. Prob. 3

$N_R V_R$ = Vol. Filtrate/min = $A_D \sqrt{\frac{2 \psi_f \Delta P N_R}{\alpha w \mu}}$ ft³/min. (Eq. 48) (Chap. 14) (4th ed.)

N_R = Rev./min.

V_R = vol. filt./rev. = ft³/rev.

A_D = Total filter area, ft²

ΔP = Pres. drop, psi

ψ_f = Fract. drum area submerged

w = lb. dry cake/ft³ filtrate

μ = fluid viscosity

α = Sp. cake resistance

From lab. data

Vol. filt./min = $\frac{0.95}{5} = 0.19$ ft³/min

$V_R = \frac{0.19}{0.4} = 0.475$ $N_R = 0.4$

$\psi_f = 0.20$

$\Delta P = \frac{9}{29.92}(14.7) = 4.42$ psi

let x = lb wet cake

$\frac{x}{1.6}$ = lb dry cake

$\frac{x}{1.6}(13) = x + 65.3$

$x = \frac{65.3 \,|\, 1.6}{\,|\, 11.4} = 9.17$ lb

Calc. of w: (Basis = 5 min)

Vol. filtrate = 0.95 ft³

lb. of filtrate = (0.95)(68.3) = 653

lb dry cake = $\frac{x}{1.6} = \frac{9.17}{1.6} = 5.73$

w = lb dry cake/ft³ filtrate

$= \frac{5.73}{0.95} = 6.03$

Basis: 1 min. + lab. data:

(Eq. 48) (Chap. 14) $0.19 = 4.15 \sqrt{\frac{2 \,|\, 4.42 \,|\, 0.2 \,|\, 0.4}{\,|\, \alpha\mu \,|\, \,|\, 6.03}}$

$$\alpha\mu = \left(\frac{4.15}{0.19}\right)^2 \left(\frac{2 \,|\, 4.42 \,|\, 0.2 \,|\, 0.4}{\,|\, \,|\, \,|\, 6.03}\right)$$

Pr. Scs. Prob. 3 (continued)

Basis: 1 min. + Design Data

100 lb. $ZnSO_4 \cdot 2.5H_2O$ in cake

(100)(20) = 2000 lb. of liquid to filter

lb. of liquid in wet cake = (100)(0.6) = 60

Wt. of wet cake lvg. filt. area = 160 lb.

Wt. of filtrate = 2000 − 60 = 1940 lb.

$$\text{Vol. of filtrate} = \frac{1940}{68.8} = 28.2\ ft^3$$

$$w = \text{lb. dry cake}/ft^3 \text{ filtrate} = \frac{100}{28.2} = 3.54$$

$N_R = 0.33$ Rev/min.

$\Delta P = 5$ psi

$\psi_F = 0.25$

(4th ed. Eq. 48) (Chap. 14)

$$28.2 = A_D\sqrt{\begin{array}{c|c|c|c} 2 & 5 & 0.25 & 0.33 \\ \hline & \alpha\mu & & 3.54 \end{array}}$$

Substitute value of $\alpha\mu$ from preceding page and solve for A_D. $A_D = \underline{\underline{437\ ft^2}}$

Answer

Vacuum Pump Power:

(4th ed. Chap. 14) (Eq. 56)

Lab. Data Vol. air/min. as ft^3 at s.c. $= \frac{8.5}{6} = \frac{A_D^2 \psi_a \Delta P}{\beta \mu_a w V_R}$

ψ_a = Fract. drum area for air suction

μ_a = air viscosity

β = sp. air cake res.

$$\frac{8.5}{5} = \begin{array}{c|c|c} \overline{4.15}^2 & 0.1 & 4.42 \\ \hline \beta\mu_a & 6.03 & 0.475 \end{array}$$

$$\beta\mu_a = \begin{array}{c|c|c|c} 5 & \overline{4.15}^2 & 0.1 & 4.42 \\ \hline 8.5 & & 6.03 & 0.475 \end{array}$$

Pr. Ses. Prob. 3 (Continued)

Design Data

$$\text{Vol. filtrate/rev.} = V_R = \frac{28.2}{0.33} = 84.6 \text{ ft}^3/\text{rev.}$$

$$\text{Vol. air/min} = \frac{\overline{437}^{\,2}}{\beta\mu_a}\Big|\frac{0.1}{3.54}\Big|\frac{5}{84.6} = 202 \text{ ft}^3 \text{ at sc.}/\text{min.}$$

$\psi_a = 0.1$

By. subst. $\beta\mu_a$ from preceding page.

$$V_1 \text{ actual or } q_{fm_1} = \frac{202}{}\Big|\frac{570}{492}\Big|\frac{14.7}{9.7}$$

$$= 355 \text{ ft}^3/\text{min at } 9.7 \text{ psi } \& \ 110°\text{F}$$

pump intake cond.

$$\text{Work} = \frac{k}{k-1}\, p_1 V_1 \left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right] (3.03\times 10^{-5}) \text{ hp.}$$

$k = \frac{c_p}{c_v} = 1.4$

p_1 = intake pres. = 9.7 (144) lb/ft.^{2}

p_2 = exhaust pres. = 14.7 (144) lb/ft.^{2}

V_1 = ft^3/min into pump = q_{fm_1} in Eq. 24 (4th ed)

$$\text{Work (hp.)} = \frac{3.03\times 10^{-5}}{}\Big|\frac{1.4}{1.4-1}\Big|\frac{(9.7)(144)}{\underset{\substack{\uparrow \\ \text{Eff}}}{0.85}}\Big|\,355\left[\left(\frac{14.7}{9.7}\right)^{\frac{0.4}{1.4}} - 1\right]$$

Motor size = $\underline{\underline{7.8 \text{ hp}}}$ Round off to $\underline{\underline{8 \text{ hp}}}$

Answer Answer.

Pr. Ses. Prob. No. 4

Determination of H.T.U.

$G_2 = 27.2$ moles/hr.

$L_2 = \frac{200 \mid 60 \mid 68.5}{7.48 \mid \quad \mid 70} = 1570$ moles/hr

$m_2 = 20 \qquad \frac{m_2 G_2}{L_2} = \frac{20 \mid 27.2}{\mid 1570} = 0.347$

HTU = 4.1 ft by linear interpolation

Determination of N.T.U

y_1^* in equil. with 0.21% PCA

$N.T.U. = N_t = \frac{y_1 - y_2}{\Delta y_m} \qquad \Delta y_1 = 0.15 - 0.042 = 0.108$

$\Delta y_2 = 0.02 - 0.002 = 0.018$

y_2^* in equil/ with 0.019% PCA

$$\Delta y_m = \frac{\Delta y_1 - \Delta y_2}{\ln \frac{\Delta y_1}{\Delta y_2}} = \frac{0.108 - 0.018}{\ln \frac{0.108}{0.018}} = 0.0503$$

$$N.T.U. = N_t = \frac{0.15 - 0.02}{0.0503} = 2.59$$

Column Ht = (2.59)(4.1) = 10.6 ft Use 11 feet

Determination of Column Diameter

Gas enters at $\frac{150}{60}$ ft³/sec. (Max. possible value would be at 35°C and 14.7 psia or 178.5 ft³/min

Gas inlet vel = $1.5/1.6$ = 0.9 ft/sec. To give D = 2.05 ft)

D = Diameter, ft: $0.785 D^2 = \frac{150 \mid}{60 \mid 0.9}$ D = 1.88 ft Use 2 ft

Column Cost

Tower jacket = (\$880)(11) = \$9680

Heads = (2)(\$750) = 1500

Packing = (0.785)(4)(11)(\$24) = 829

or 12,009

Total cap. expend. = \$12,009 + \$80,000 = \$92,009

% Return on investment = $\left(\frac{40,000}{92,009}\right)100$ = 43.5%

Answer

Pr. Ses. Prob. No. 5

% Cl_2	Net Annual Savings, $	% Return on Capital Investment
0.2	120,000	32.6
1.0	128,000	42.0
2.0	124,000	45.6
5.0	64,000	30.2
10.0	-48,000	-35.4

For Part 4

% Cl_2	Inv	Net Svg.	Ret	%
2.5	260,000	116,000	8000	$\frac{8000}{120}$ Yes
2.0	272,000	124,000		
1.5	284,000	124,000	0	$\frac{0}{120}$ No
1.0	304,000	128,000	4000	$\frac{4000}{200}$ No
0.5		124,000	No	

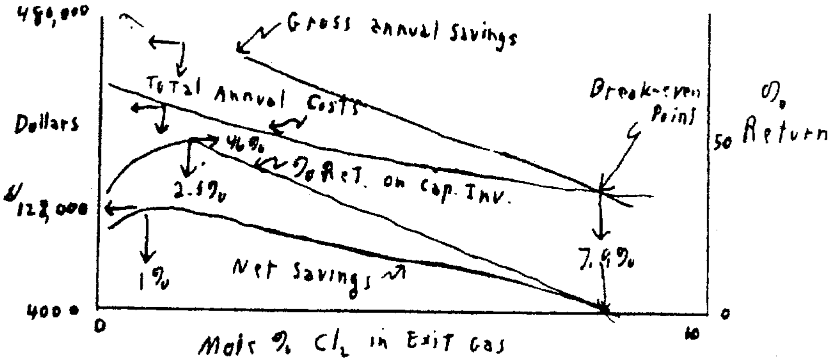
From Plot

1. % Cl_2 at break even point = 7.9 %
Answer

2. % Cl_2 at Max. Net Annual Savings = 1 %
Max. Net Annual Savings = $128,000 // Answer

3. % Cl_2 at Max. % Return = 2.5 %
Max. % Return = 46 % // Answer

4. See comparison above in right upper corner:
It is close in range of 1.5 to 2 %
Recommend 2 % Cl_2 with $272,000 investment
Answer



Pr. Ses. Prob. 6

D = O.D. of lagging in inches

t_s = T° of outside lagging surface in °F

With insulation:

Initial inv. for insul. $= 0.25\, D^{1.3}(3)(5280) = 3960 D^{1.3}$

Yearly Deprec. $= \frac{3960 D^{1.3}}{20} = 198 D^{1.3}$ \$/yr.

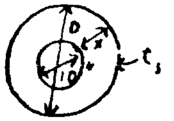
Total cost of equip/yr. $= 198\, D^{1.3} + 3960 D^{1.3}\left(\frac{10+2}{100}\right) = 673.2\, D^{1.3}$ \$/yr.

$$Q = UA\,\Delta t = \frac{k\, A_{avg}\, \Delta t_L}{x} = (h_c + h_r)(A_o)\,\Delta t_s$$

x = ins. thickness $= \frac{D-10}{2}$ inches

$$A_{avg} = \frac{D-10}{\ln \frac{D}{10}}\, \frac{\pi}{12}(3)(5280)$$

or assume $\frac{A_o}{A_i} < 2.0$; then $A_{avg} = \left(\frac{D+10}{2}\right)\frac{\pi}{12}(3)(5280)$



$A_o = \frac{D}{12}\pi(3)(5280)$

$A_i = \frac{10}{12}\pi(3)(5280)$

$\Delta t_L = 250 - t_s$

$\Delta t_s = t_s - 45$

$x = \frac{D-10}{2}$ inches

$h_c + h_r = 0.42\left(\frac{\Delta t_s}{D}\right)^{0.25} + 1.2$

$Q_{\text{Conv. \& Rad.}}$ (as BTU/hr) =

$$\left[1.2 + 0.42\left(\frac{t_s-45}{D}\right)^{0.25}\right]\frac{D}{12}\pi(3)(5280)(t_s-45)$$

Eq. Ⓐ

$Q_{cond.}$ (as BTU/hr) =

$$(0.04)\frac{\pi}{12}(3)(5280)\frac{D-10}{\ln\frac{D}{10}}\,\frac{(250-t_s)}{D-10}(2)(12)$$

Eq. Ⓑ

Eq. Ⓒ

Ⓐ = Ⓑ ∴

$$\left[1.2 + 0.42\left(\frac{t_s-45}{D}\right)^{0.25}\right]D(t_s-45) = (0.04)\frac{D-10}{\ln\frac{D}{10}}\,\frac{(250-t_s)}{(D-10)}(2)(12)$$

Pr. Ses Prob 6 (Continued)

Without Insulation $Q'(BTU/hr.) = (h_c + h_r) A_o (\Delta t_s)$

$A_o = \pi \frac{10}{12}(3)(5280)$ $\Delta t_s = 250-45$ $h_c = 0.42\left(\frac{250-45}{10}\right)^{0.25}$

$h_c = 0.895 \frac{BTU}{hr\ ft^2\ °F}$

$Q' = (0.895 + 1.2)\pi \frac{10}{12}(3)(5280)(205) = 1.78 \times 10^7 \frac{BTU}{hr}$

Saving on steam with ins. as \$/yr. =

$$\left[1.78\times10^7 - (0.04)\frac{\pi}{12}(3)(5280)\frac{D-10}{\ln\frac{D}{10}}\left(\frac{250-t_s}{D-10}\right)(2)(12)\right]\overbrace{(24)\left(\frac{365}{\underset{\nearrow}{945.5}}\right)\left(\frac{1.30}{1000}\right)}^{0.01201}$$

ht. of evap of H_2O at 250°F

Net savings with Ins. as \$/yr = Steam savings - Costs

Return on Investment = $\frac{\text{Net savings/yr}}{\text{Total Inv.}} = 0.50 =$

$$\frac{\left[1.78\times10^7 - (0.04)\frac{\pi}{12}(3)(5280)\frac{D-10}{\ln\frac{D}{10}}\left(\frac{250-t_s}{D-10}\right)(2)(12)\right](0.01208) - 673.2D^{1.3}}{3960D^{1.3}}$$

Eq. Ⓓ ↗

METHOD FOR COMPLETING SOLUTION

Assume a value of D (or t_s) and calculate the value of t_s (or D) from eq. C. Put these values of D and t_s into eq. D and check the assumption. Repeat until eq. D is satisfied

Lagging thickness = $\frac{D-10}{2}$ inches

Answer

Pr. Ses. Prob. No. 7

At start, moles isopropanol consumed/hr. = $\frac{76.1}{0.98}$

N = Moles isoprop. fed to converter/hr = $\frac{76.1}{0.98\,\alpha} = \frac{77.6}{\alpha}$ (Keep constant)

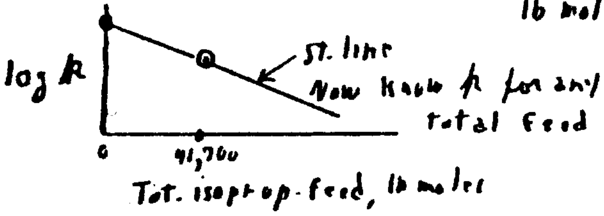
At start, $k = 0.3 = \frac{0.000254 \;\big|\; 77.6 \;\big|\; 572+460}{250 \;\big|\; \alpha \;\big|\;}\left[2.46 \ln\frac{1}{1-\alpha} - \alpha\right]$

Solve for α ($\alpha = 0.78$)

Therefore, can solve for $N = \frac{77.6}{\alpha}$ (= 99.7 moles/hr)

k is 0.15 sec^{-1} when total feed = $\frac{10{,}000 \;\big|\; 250}{\;\big|\; 60.1} = 41{,}700$ lb moles

Fig. 1 plot



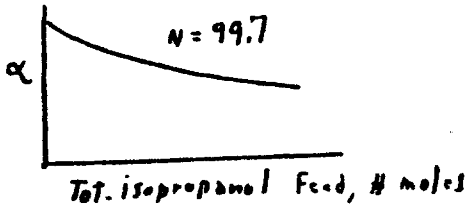
Need α at various total feeds

$$k = \frac{0.000254 \;\big|\; \overset{N}{99.7} \;\big|\; 1032}{\;\big|\; 250 \;\big|\;}\left[2.46 \ln\frac{1}{1-\alpha} - \alpha\right]$$

Fig. 2 plot

From Fig. 1 and Fig. 2, plot

Fig. 3



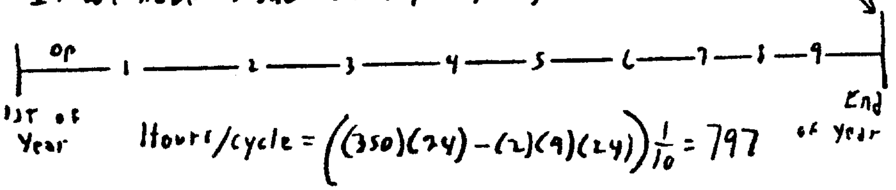
Pr. Ses. Prob. 7 (Continued)

(Neglect down time)

Get moles of isopropanol feed per cycle

If we have 9 shutdowns per year,

Throw away catalyst

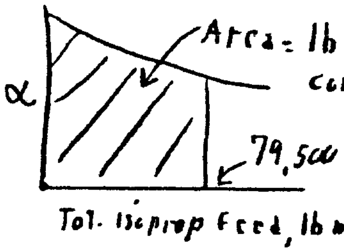


$$\text{Hours/cycle} = \left((350)(24) - (2)(9)(24)\right)\frac{1}{10} = 797$$

Total isopropanol fed/cycle = 797(99.7) = 79,500 lb moles

$$\alpha = \frac{d\,(\text{isoprop. cons.})}{d\,(\text{isoprop. sup.})} \quad \text{or} \quad \text{isoprop cons.} = \int_0^{79,500} \alpha\, d(\text{isoprop. sup})$$

From Fig. 3



Area = lb moles isoprop. consumed per cycle = 50,500

Acetone produced in lb per year =

(moles isoprop cons./cycle)(efficiency)(cycles/yr)(M.W. of Acetone)

= (50,500) (0.98) (10) (58.1) = 28,800,000 lb/yr

Pr. Ses. Prob. No. 8

Get amount of acetone prod. at max. prod rate $(k = 0.3\ \text{sec}^{-1})$

$$k = \frac{0.000254}{V} NT \left[2.46 \ln \frac{1}{1-\alpha} - \alpha\right] \qquad N = 97 \quad V = 250 \quad T = 1032^\circ R \quad k = 0.3$$

Solve for α: $\alpha = \underline{0.78}$

lb acetone delivered from top of dist col. as product =

$$(97)(0.78)(0.98)(0.98)(58.1) = 4220\ \text{lb/hr.}$$

Necessary column cross sect. area = $\frac{4220}{1000}(6.4) = 27.0\ \text{ft}^2$

Actual column cross sect. area = $0.785(6)^2 = 28.3\ \text{ft}^2$

The present column is satisfactory

The present condenser is satisfactory

Calandria Calc.

ht. of cond. of steam at 64.7 psia

$$q = \frac{4220}{1000}(1100)(911.8) = UA\Delta t \ \ \text{Btu/hr}$$

$U = 250\ \text{BTU/hr ft}^2\ {}^\circ\text{F}$ $\Delta t = 90^\circ\text{F}$

Calculate value of $A = 188\ \text{ft}^2$

Available area is $104\ \text{ft}^2$; so supply new calandria

If Use 20% safety factor, Area = $188 \,|\, 1.2 = 226\ \text{ft}^2$

Therefore, Area of at least $200\ \text{ft}^2$ should be provided.

Cost = (\$160) (200) = \$32,600 at least

Answer

Pr. Scs. Prob. No. 9

	CHARGE STOCK				OVHD		BTMS	
Comp.	Mol %	M.W.	Avg M.W	Mols/Hr	Mols/Hr	Mol %	Mols/Hr	Mol %
C_2''	0.1	28	0.03	0.8	0.8	0.2		
C_2	1.2	30	0.36	10.0	10.0	1.9		
H_2S	2.1	34	0.71	17.5	17.5	3.3		
C_3''	16.3	42	6.85	135.7	135.7	26.3		
C_3	6.9	44	3.03	57.5	57.5	11.1		
IC_4''	6.5	56	3.64	54.2	54.2	10.5		
NC_4''	14.3	56	8.00	119.2	119.2	23.0		
IC_4	10.8	58	6.26	90.0	90.0/484.9	17.4		
LK → NC_4	3.9	58	2.26	32.4	25.0	4.8	7.4	2.4
C_5''	11.9	70	8.33	99.1	4.8	0.9	94.3	29.9
HK → IC_5	9.7	72	6.98	80.8	3.0	0.6	77.8	24.7
NC_5	2.3	72	1.65	19.2			19.2	6.1
C_6	11.8	86	10.14	98.3			98.3	31.1
C_7	2.1	96	2.02	17.5			17.5	5.5
C_8	0.1	112	0.11	0.8			0.8/135.8	0.3
	100.0	60.4≈	60.35	833.0	517.7	100	315.3	100

% OVHD =

$$\frac{517.7}{833}(100) = 62.2\%$$

Get Sp. Gr. 60/60

$$°API = \frac{141.5}{Sp.Gr.\ 60/60} - 131.5 = 100.3$$

Sp. Gr. = 0.613

$$\text{Total moles/hr} = \frac{5620 \mid 42 \mid 0.613 \mid 62.4}{\mid 7.48 \mid 24 \mid 604} = \underline{833}$$

Get No. of Trays & Amount of Reflux From Card A

About the same case: Card A has 66.2% OVHD. We have 62.2% OVHD and poorer C_5-C_4 separation. Fewer plates and lower reflux ratio could be used.

Use 27 Trays and return 115 mol% Feed as Reflux

Pr. Ses. Prob. #9 (Continued)

Distillate = 517.7 moles/hr Reflux = $\frac{115}{100}$ (833) = 957 moles/hr

Assume feed near B.P.; so $V = \bar{V}$ and diameter is set for $V = 517.7 + 957 = 1474.7$ moles/hr

At one Atm. pres. = assume MAV = 3 ft/sec.

At 165 psia & 85% of MAV, the MAV (superficial) =

$$3\sqrt{\frac{14.7}{165}}(.85) = 0.77 \text{ ft/sec.}$$

Determine Diameter at point of higest Temp.

Thus, get Temp. at reboiler. Use $y = kx$ and Card B.

With x values for BTMS as given and k values on card B, get y for each component at trial and error temperature until $\Sigma y = 1.0$.

Result is; Temp. = 282°F

Thus
$$0.77 = \begin{array}{c|c|c|c|c|c|c} 1474.7 & 359 & 14.7 & 460+282 & & & \\ \hline & & 165 & 492 & 3600 & 0.785 & D^2 \end{array}$$

D = 5.73

For Reboiler Area - Use Card C $U = 130$ BTU/hr ft² °F is safe

$$q = (1474.7)(5000)(1.8) = (130)A(\underset{\uparrow}{401}-282)$$

t° of cond. of steam at 250 psia

A = 860 ft²

Final Results Use column with 27 trays ± 10%

Use column diameter of 5.73 ft ± 10%

Use Reboiler area of 860 ft² ± 10%

Answer

Pr. Ses. Prob. No. 10

Determine cost of Prod. HCHO

Basis: 1 operating day

$\text{F.C.} = \frac{1,140,000}{350}\Big|\frac{15}{100} = \$\,489$

$\text{Misc.} = \frac{360,000}{350} = \$\,1029$

$CH_3OH\ \text{cost} = (57,600)(0.091) = \$\,5242$

$\text{Utilities cost} = x = \$\,356$

$\text{Total cost}\ \$\,7116$

$CH_3OH + \frac{1}{2}O_2 \rightarrow HCHO + H_2O$

$\underset{32}{CH_3OH} \rightarrow \underset{30}{HCHO} + H_2$

lb. Methanol necess =

$\frac{100,000}{}\Big|\frac{37.2}{100}\Big|\frac{32}{30}\Big|\frac{}{6.8}$

$+ \frac{100,000}{}\Big|\frac{8}{100} = 57,600$

If utilities cost = x

$\frac{x}{x + 489 + 1029 + 5242} = 0.05$

Total income from HCHO only $= (100,000)(0.100) = \$\,10,000$

Present profit/yr $= (10,000 - 7116)(350) = \$\,\underline{\underline{1,009,000}}$

Answer 1

Situation with P.E. Plant in use

Basis: One operating day

$\underset{56.1}{CaO} + H_2O \rightarrow \underset{74.1}{Ca(OH)_2}$

$\underset{74.1}{Ca(OH)_2} + 2\,\underset{88}{CH_3CHO} + \underset{240}{8\,HCHO} \rightarrow 2\,\underset{272}{C(CH_2OH)_4} + \underset{130}{Ca(COOH)_2}$

Final P.E. Prod. = 6000 lb.

P.E. Prod. in reaction $= \frac{6000}{0.7} = 8570$ lb.

lb $Ca(OH)_2$ necess. $= \frac{74.1}{272}\Big|\frac{8570}{} = 2330$ lb.

lb CH_3CHO necess. $= \frac{88}{272}\Big|\frac{8570}{} = 2780$ lb. $\simeq \frac{2780}{}\Big|\frac{0.46}{} = \$\,1279$

lb HCHO necess. $= \frac{240}{272}\Big|\frac{8570}{0.372} = 20,400$ lb. $\simeq \frac{20,400}{}\Big|\frac{0.07116}{} = 1452$

lb CaO necess. $= \frac{2330}{}\Big|\frac{56.1}{74.1} = 1770$ lb $\simeq \frac{1765}{}\Big|\frac{40}{2000} = 35$

Total raw mtl. cost $= \$\,2766$

Pr. Ses. Prob. No. 10 (Continued)

All other costs = $\frac{0.4 \mid 500{,}000}{\mid 350} = \$571$

Total cost = 571 + 2766 = \$3337 per day

Cost per lb. = $\frac{3337}{6000} = \$0.556$ per lb

Total profit per year with P.E. plant operating =

$$\frac{0.72 - 0.556 \mid 6000 \mid 350}{\mid \qquad \mid} +$$

$$+ \frac{0.100 - 0.07116 \mid 100{,}000 - 20{,}400 \mid 350}{\mid \qquad \mid}$$

= \$1,147,500 Total Yearly Profit

Answer 2

Yearly return on P.E. Investment = 1,147,500 − 1,009,600 = \$138,500

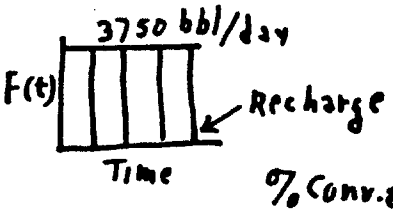
% return on P.E. Investment $\frac{138{,}500}{500{,}000}(100) = 27.7\%$

Answer 3

The % return does not reach the required 30%. Therefore, the investment should not be made in the P.E. plant. NO

Answer 4

Pr. Ses. Prob. 11



$F = \frac{15{,}000}{4}\left|\frac{\,}{1000}\right. = 3.75$

or 3750 Barrels/day

% Conv. at cat. age A = 93.75% by problem statement

If % conversion at Time 0 (Cat. Age Factor = A = 0) is 97.66%, then $\frac{\%\text{ Conv. at Cat. Age } A}{\%\text{ Conv. at Cat. Age } 0} = \frac{93.75}{97.66} = 0.9600$

and, by table in Problem, A = Cat. age factor = 50

$$\log \frac{\Delta P}{F^2} = 0.03179A - 0.699 = 0.03179(50) - 0.699$$

$$= 0.8905 = \log \frac{\Delta P}{(3.75)^2}$$

$\Delta P = 109.3$ psi at 93.75% Conv. Answer

$$A = \frac{\left[\int_0^D F(t)\,dt\right] \times \left[\text{conv. factor to give gal. of polymer}\right]_{a.a.}}{\text{Pounds of catalyst}}$$

Arith. Avg. (a.a.) conv. factor to give gal. of polymer

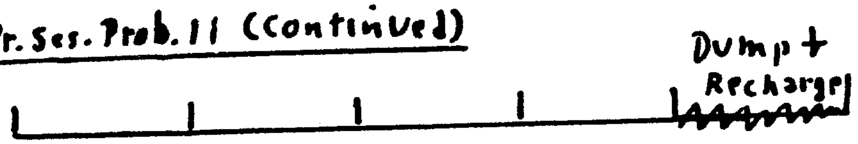
$$= 42\,\frac{\text{gal.}}{\text{bbl}} \times 0.4\,\frac{\text{bbl propylene in feed}}{\text{bbl of feed}} \times 0.715\,\frac{\text{bbl of polymer prod.}}{\text{bbl of propylene conv.}}$$

$$\times \frac{93.75 + 97.66}{2 \times 100}\left[\frac{\text{bbl of propylene conv.}}{\text{bbl of propylene in feed}}\right]_{a.a.} = 11.50$$

$$A = 50 = \left[\int_0^D F(t)\,dt\right] \times \frac{11.50}{20{,}000} = \frac{15{,}000}{4}\,D\,\frac{11.50}{20{,}000}$$

$D = \frac{50}{\,}\left|\frac{20{,}000}{15{,}000}\right|\frac{4}{11.5} = 23.19$ Days each reactor is on stream
Answer

Pr. Ses. Prob. 11 (Continued)

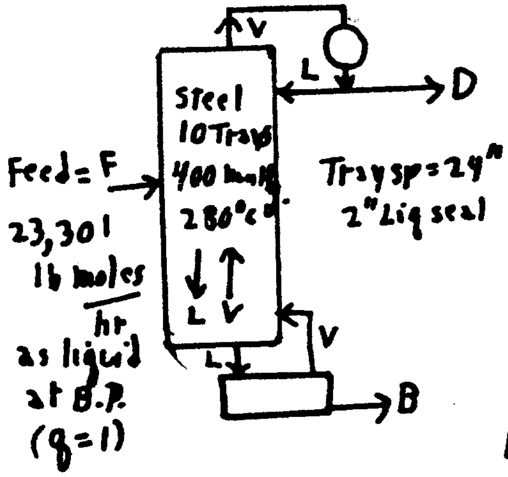


Dump + Recharge time = $\frac{23.19}{4}$ = 5.80 Days

Time each unit should come on stream = Days between each new recharged unit coming on stream = 5.8 Days

Answer

Pr. Ses. Prob. 12



F, D, B, V, + L = lb moles/hr

Assume const. T° + Pres. + ideal gas + adiabatic + const. molal overflow + q = 1.

Gas m.w. = 154.2

Liq. dens = 0.72×62.4 = 44.9 lb/ft³

D contains 98% of the biphenyl entering

B contains 2% of the biphenyl entering and all below that

D contains all entering above biphenyl.

Pr. Ses. Prob. 12 (Continued)

Material Balances

$\frac{L}{D} = R.R. = 8 \qquad L = 8D$

$V = L + D = 9D$

$D = 0.488 + 1.599 + 0.98(16.835) = 18.5853 \ \frac{\text{lb moles}}{\text{hr}}$

$V = 9D = (9)(18.5853) = 167.2677 \ \frac{\text{lb moles}}{\text{hr}}$

$$\rho_G = \frac{m}{Vol} = \frac{PM}{RT} = \frac{154.2}{359}\Bigg|\frac{273}{280+273}\Bigg|\frac{400}{760} = 0.112 \ \frac{lb}{Ft^3}$$

$\rho_L = 44.9 \ lb/FT^3$

Using Eq. 1 (Chap. 16) & Fig. 16-6; $K_V = 0.30$ For 24" plate spacing

$$V_m = K_V\sqrt{\frac{\rho_L - \rho_G}{\rho_G}} = 0.30\sqrt{\frac{44.9 - 0.112}{0.112}} = 6.0 \ Ft/sec$$

Actual superficial $V_m = (6.0)(0.85) = 5.1$ Ft/sec

Let D_F = Column diameter in feet

$$\frac{167.2677}{}\Bigg|\frac{}{3600}\Bigg|\frac{359}{}\Bigg|\frac{760}{400}\Bigg|\frac{553}{273}\Bigg|\frac{}{0.785 D_F^2} = 5.1$$

$D_F = 4.0$ FT

By Fig. 16-28, Cost per plate = \$4100 in Jan., 1990.
This includes installation and auxiliaries.

Cost of Tower installed with auxiliaries (no reboiler or condenser) = (\$4100)(10) = \$41,000

Answer

Pr. Ses. Prob. 12 (Continued)

Using Eq. 3 (Chap 16) + Fig. 16-7; (4th ed.)

$$V_m' = K_v' \left(\frac{\sigma}{20}\right)^{0.2} \sqrt{\frac{\rho_L-\rho_G}{\rho_G}}$$

$$\frac{L}{G}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} = \frac{8 \times 18.5}{166.6}\left(\frac{0.112}{44.9}\right)^{0.5} = 0.044$$

By Fig. 16-7, $K_v' = 0.36$

Assume no correction factors needed and $V_m = 0.95 V_m'$ as in Example 1 of Chap. 16 (4th ed.)

$$V_m = (0.36)\left(\frac{20}{20}\right)^{0.2} \sqrt{\frac{44.9-0.112}{0.112}}\,(0.95) = 6.8 \ \frac{ft}{sec}$$

Actual superficial $V_m = (6.8)(0.85) = 5.8$ ft/sec

$$\frac{167.2677}{3600} \cdot 359 \cdot \frac{760}{400} \cdot \frac{553}{273} \cdot \frac{1}{0.785 D_F^2} = 5.8$$

$D_F = 3.8$ ft

By Fig. 16-28, cost per plate = \$3900 in Jan., 1990.

Cost of Tower installed with auxiliaries (no reboiler or condenser) = (\$3900)(10) = \$39,000

Answer

Pr. Ses. Prob. 13

Case of Single back-mix reactor operated Isothermally at 800°F

$$\frac{W}{Q} = C_{S_0}\left(\frac{X_{SF} - X_{Si}}{-r_S}\right)$$

$$-r_S = k_S A_S \frac{C_S^2 P_H}{C_{S_0}(1 + K_{HS} P_{HS})} \text{ lb moles/hr-lb cat.}$$

$C_S = C_{S_0}(1 - X_{SF})$

$P_H = P_{H_F} = 2075$ psia

$P_{HS} = P_{HS_F} = 4.4$ psia

$K_{HS} = 0.162$ psia^{-1}

$C_{S_0} = 0.0213$ lb mole/ft^3

$X_{Si} = 0$

$X_{SF} = 0.74$

$k_S = 7.405 \times 10^{-4}$ ft^3/hr-lb cat-psia

$A_S = 1.25$

Cat. Density = 42 lb/ft^3

4.2 Bbl. of Prod oil/ton coal

At 50,000 Bbl of prod. oil/day, need 50,000/4.2 tons coal/day

Therefore,

$$Q = \frac{50,000}{4.2} \times \frac{1}{24} \times 127.34 \frac{\text{ft}^3 \text{ slurry}}{\text{Ton coal}} = 63,300 \text{ ft}^3 \text{ slurry/hr}$$

$$-r_S = k_S A_S \frac{C_{S_0}(1 - X_{S_F})^2 P_{HF}}{(1 + K_{HS} P_{HS_F})}$$

$$-r_S = \frac{7.405 \times 10^{-4} \;\big|\; 1.25 \;\big|\; 0.0213 \;\big|}{\big|\qquad\big|\qquad\big|}$$

$$\frac{(1-0.74)^2 \;\big|\; 2075}{1 + (0.162)(4.4) \;\big|} =$$

$$= 0.001625$$

$$W = \text{lb cat. charge} = Q C_{S_0} \frac{X_{S_F} - X_{Si}}{-r_S}$$

$$W = \frac{63,300 \;\big|\; 0.0213 \;\big|\; (0.74 - 0)}{\big|\; 0.001625 \;\big|}$$

$$= 613,991$$

Vol. of CSTR needed if no extra space other than for catalyst = $\frac{613,991}{42} = 14,600$ ft^3 Answer

Pr. Ses. Prob. 13 (Continued)

Case of Ideal, Single, Plug-flow Reactor Operated Isothermally at 800°F

$$\frac{W}{Q} = C_{S_0}\int_{x_{S_i}}^{x_{S_F}} \frac{dx_S}{-r_S} \qquad -r_S = k_S A_S \frac{C_S^2\, p_H}{C_{S_0}(1+K_{HS}\,p_{HS})}$$

$$-r_S = \frac{k_S A_S\, p_{H_{avg}}\, C_{S_0}}{\underbrace{(1+K_{HS}\,p_{HS})_{Avg.}}_{\text{constant by assumption}}}(1-x_S)^2 \ \frac{\text{lb moles}}{\text{hr-lb cat.}}$$

$$= \frac{(1+0)+[1+(0.162)(4.4)]}{2} = 1.3564$$

$C_{S_0} = 0.0213$ lb moles/ft^3

$p_{H_{avg}} = \frac{2125+2075}{2} = 2100$ psia

$p_{HS_0} = 0$

$K_{HS} = 0.162$ psia^{-1}

$x_{S_i} = 0$

$x_{S_f} = 0.74$

$C_{S_0} = 0.0213$ lb mole/ft^3

$k_S = 7.405\times10^{-4}$ ft^3/hr-lb cat.-psia

$A_S = 1.25$

$Q = 63,300$ ft^3 slurry/hr

Cat. density = 42 lb/ft^3

Pr. Ser. Prob. 13 (Continued)

$$-r_s = \frac{7.405\times10^{-4}}{}\bigg|\frac{1.25}{}\bigg|\frac{2100}{}\bigg|\frac{0.0213}{1.3564}\left(1-x_s\right)^2$$

$$-r_s = 0.03057\left(1-x_s\right)^2$$

$$W = \frac{Q\,C_{s0}}{0.03057}\int_0^{x_{s_f}}\frac{dx_s}{\left(1-x_s\right)^2}$$

$$\int_0^{x_{s_f}}\frac{dx_s}{(1-x_s)^2} = \left.\frac{1}{1-x_s}\right]_0^{x_{s_f}=0.74} = 2.85$$

$$W = \frac{63,300}{}\bigg|\frac{0.0213}{0.03057}\bigg|\frac{2.85}{} = 125,699 \text{ lb. cat. needed}$$

Vol. of plug-flow reactor needed if no extra space other than for catalyst =

$$\frac{125,699}{42} = 3,000 \text{ ft}^3$$

Answer

Versus 14,600 ft^3 needed for CSTR

Pr. Ses. Prob. 14

$$\frac{1700 \text{ m}^3 \text{ alkylate/day}}{1.72 \ \dfrac{\text{m}^3 \text{ alkylate}}{\text{m}^3 \text{ butylene consumed}}} = 988.37 \text{ m}^3 \text{ Butylene consumed/day}$$

$$\frac{988.37}{0.74} = 1336 \text{ m}^3 \text{ Butylene Feed/day}$$

$$918.37 \text{ m}^3 \text{ Butylene consumed/day} \times 1.10 \ \frac{\text{m}^3 \text{ isobutane cons.}}{\text{m}^3 \text{ butylene cons.}}$$

$$= 1087.2 \text{ m}^3 \text{ isobutane consumed/day}$$

$$\underbrace{(1336)(0.17)}_{\text{isobutane from butylene feed}} + \underbrace{(0.9)(\text{Isobutane feed/day})}_{\text{isobutane from isobutane feed}} = 1087.2$$

Isobutane feed/day = 956 m^3/day

Answer (a)

$$\text{Reactor feed} = \underbrace{1336 + 956}_{2291.36} + \underbrace{\text{Recycle}}_{R \text{ as m}^3/\text{day}}$$

Volume change of flow stream through reactor =

$$\underbrace{(1.72)(988.37)}_{\text{Inc. in vol. due to Butylene conv. to alkylate}} - \underbrace{(1.1)(988.37)}_{\text{Loss in vol due to isobutane conv. to alkylate}} - \underbrace{988.37}_{\text{Loss in vol due to Usage of butylene}} = -375.6 \text{ m}^3$$

Pr. Ses. Prob 14 (Continued)

P = Reactor product as m^3/day $= 2291.36 + R - 375.6$

$= 1915.76 + R$

Reactor product is 75 vol-% isobutane, and, since all isobutane coming out of reactor must be in recycle stream R (all exit streams are assumed to be pure),

$R = 0.75P$ and $P = 1915.76 + 0.75P$

$P = \frac{1915.76}{0.25} = 7663.12$ m^3 prod./day

R = amout of isobutane recycled =

$(0.75)(7663.12) = 5750$ m^3/day

Answer

Pr. Ses. Prob. 15

See solution to 1970 AIChE Stud Cont. Prob. in Fall, 1970 AIChE Stud. Members Bul. - P. 75 - Table

Part A: 20% Cont. nom. int. rate of return after taxes required on any investment

$\text{T.C.I.} \overset{\checkmark}{=} 300{,}000 = \text{F.C.I.} + \text{W.C.I.} = \text{F.C.I.} + 0.1\ \text{F.C.I.} = 1.1\ \text{F.C.I.}$

$\text{F.C.I.} = 300{,}000/1.1 \overset{\checkmark}{\cong} 273{,}000.$ $\quad \text{W.C.I.} \overset{\checkmark}{=} 27{,}000$ ↑ round off

Annual Depr. cost $= \dfrac{273{,}000 - 0}{10} \overset{\checkmark}{\cong} 27{,}000$ ↑ round-off

Annual cost without depreciation before taxes $= 758{,}000 - 27{,}000 + 0.05y = 731{,}000 + 0.05y$

+27,000 startup in 1st yr. only

y = annual income before taxes

A. Taxable profit per normal year $= \left[y - \left(\underset{\text{Incl. depr.}}{758{,}000} + 0.05y\right)\right]$

↗ 48% of this is tax cost

B. Income taxes per normal year $= 0.48\left[y - \left(758{,}000 + 0.05y\right)\right]$

In-an-instant at end of year,

Cash flow to project after taxes in normal year $=$

$0.52\left[y - \left(758{,}000 + 0.05y\right)\right] + 27{,}000$

Cash flow to project after taxes in 1st year $=$

$0.52\left[y - \left(758{,}000 + 27000 + 0.05y\right)\right] + 27{,}000$

Cash flow to project after taxes in 10th year $=$

$0.52\left[y - \left(758{,}000 + 0.05y\right)\right] + \underset{\text{deprec.}}{27{,}000} + \underset{\text{W.C.I.}}{27{,}000}$

Pr. Ses. Prob. 15 (Continued)

YEAR	Cost-Incl. Depr.	Inc. before Taxes	Cash Flow = A to Project after Taxes	r = 20% Discount factor (Fa). Table 5 Chap 7		Pres. Value = Fa(A)
				100 r n	Fa	
0	(300,000)					
1	785,000 + 0.05Y	Y	$0.52[Y-(785,000)+0.05Y)]$ $+27,000$	20	0.8187	
2	758,000 + 0.05Y		$0.52[Y-(758,000+0.05Y)]$ $+27,000$	40	0.6703	
3	"		"	60	0.5488	
4	"		"	80	0.4493	
5	"		"	100	0.3679	
6	"		"	120	0.3012	
7	"		"	140	0.2466	
8	"		"	160	0.2019	
9	"		"	180	0.1653	
10	"	↓	$0.52[Y-(758,000+0.05Y)]$ $+27,000+27,000$	200	0.1353	
					$\Sigma F_a = 3.9053$	ΣP.V.

Profits Present Value after Taxes = Σ P.V. =

$$(3.9053)(0.52)[0.95Y-758,000)-(0.52)(27,000)(0.8187)+(3.9053)(27,000)$$
$$+(0.1353)(27,000)$$

Profits Present Value after Taxes $= 300,000 = 1.929Y - 1,441,712$

$$Y = \frac{300,000 + 1,441,712}{1.929} = \$902,909$$

$$\text{Price} = \frac{902,909}{10 \times 10^6} = \underline{\underline{0.0903}} \text{ per lb. of prod.}$$

Answer A

Note: In 1970 AIChE Stud. Contest Solution, error was made in yr 2 on by including deprec as a cost and did not correct for income-tax effect correctly, Also had error in year 10 cash flow by taking 0.52 of W.C.I.

Pr. Ses. Prob. 15 (Continued)

Part B, 20% finite eff. end-of-year rate of return after Taxes required on any investment

Year	Cost-Incl. Depr.	Inc. before taxes	Cash flow = A to Project after Taxes	i = 20% Discount factor = D.F. = $\frac{1}{(1+i)^n}$ Table 1 Chap. 7	Pres. Value = (D.F.)(
0	(300,000)				
1	(785,000 + 0.05Y)	Y	$0.52[Y-(785,000+0.05Y)] + 27,000$	$\frac{1}{1.2} = 0.833$	
2	(758,000 + 0.05Y)	↓	$0.52[Y-(758,000+0.05Y)] + 27,000$	$\frac{1}{1.44} = 0.6944$	
3	"	↓	"	$\frac{1}{1.728} = 0.5787$	
4	"	↓	"	$\frac{1}{2.0736} = 0.4823$	
5	"	↓	"	$\frac{1}{2.4883} = 0.4019$	
6	"	↓	"	$\frac{1}{2.9860} = 0.3349$	
7	"	↓	"	$\frac{1}{3.5832} = 0.2791$	
8	"	↓	"	$\frac{1}{4.2998} = 0.2326$	
9	"	↓	"	$\frac{1}{5.1598} = 0.1938$	
10	"	↓	$0.52[Y-(758,000+0.05Y)] + 27,000 + 27,000$	$\frac{1}{6.1917} = 0.1615$	
				$\Sigma D.F. = 4.1922$	$\Sigma P.V.$

Profits Present Value after taxes = $\Sigma P.V. = 300,000$

$$= (4.1922)(0.52)(0.95Y - 758,000) - (0.52)(27,000)(0.833) + (4.1922)(27,000) + (0.1615)(27,000)$$

$$= 2.0709Y - 1,546,534 = 300,000$$

$$Y = \frac{300,000 + 1,546,534}{2.0709} = 891,662$$

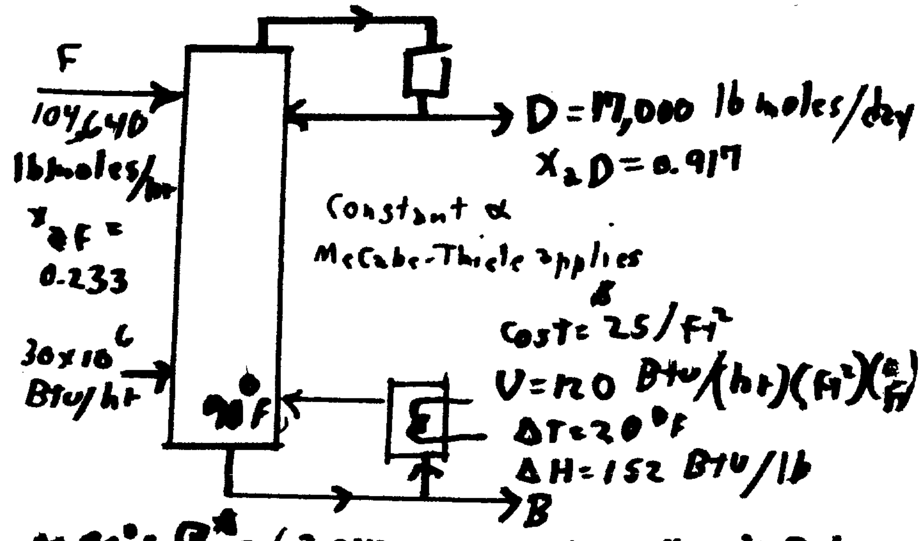
$$\text{Price} = \frac{891,662}{10 \times 10^6} = \underline{\underline{0.0892}} \text{ per lb. of Prod}$$

Answer B

Pr. Ses. Prob. 16

Basis: 1 day

a = i-butane
b = n-butane



At 90°F $P_a^* = 62$ psia

$P_b^* = 44$ psia

?= Cost of reboiler at min. R.R.

Overall Mt'l Balance

$F = D + B$

$B = 104,640 - 17,000$
$= 87,640$ lb moles

$x_{aF} F = x_{aD} D + x_{aB} B$

$0.233 \times 104,640 = 0.917 \times 17000 + x_{aB} \times 87,640$

$x_{aB} = 0.1003$

Get α at Bottom temp. of 90°F and constant α:

$p_a = P_a^* x_a = 62 \times 0.1003 = 6.2185$ psia

$p_b = P_b^* x_b = 44 \times 0.8997 = 39.5868$ psia

$P_T = 45.8053$ psia

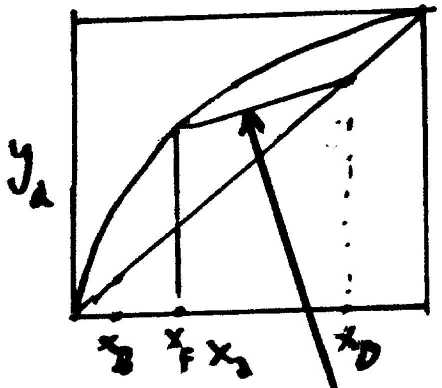
Pr. Sec. Prob. 16 (Continued)

$$y_a = \frac{P_a}{P_T} = \frac{6.2185}{45.8053} = 0.136$$

$$y_b = \frac{P_b}{P_T} = \frac{39.5865}{45.8053} = 0.864$$

$$\alpha = \frac{y_a}{y_b}\,\frac{x_b}{x_a} = \frac{0.136}{0.864}\,\frac{0.8997}{0.1003} = 1.41$$

McCabe Thiele Diagram



For Feed, $x_{aF} = 0.233$

$x_{bF} = 0.767$

$y_{bF} = 1 - y_{aF}$

$$\alpha = 1.41 = \frac{y_{aF}}{1-y_{aF}}\,\frac{0.767}{0.233}$$

$y_{aF} = 0.300$

Enriching line. For min. R.R.

$$\text{Slope} = \frac{0.917 - 0.300}{0.917 - 0.233} = 0.902 = \frac{L}{V}$$

$$V = L + D = 0.902V + 17{,}000$$

$$V = \frac{17{,}000}{0.098} = 173{,}469 \text{ lb moles/day} \times \frac{1}{24} = 7228 \ \frac{\text{lb moles}}{\text{hr}}$$

$V = \bar{V}$

Pr. Ses. Prob 16 (continued)

Reboiler heat needed $= (\bar{V})(\Delta H)(M.W.) -$

$$= (7228)(152)(58.12) - \overset{30\times10^{6}\ \frac{\text{BTU}}{\text{hr}}}{30\times10^{6}}$$

$$= 33.85\times10^{6}\ \text{BTU/hr}$$

$$Q = UA\Delta T$$

$$A = \frac{Q}{U\Delta T} = \frac{33.85\times10^{6}}{(120)(20)} = 14{,}104\ \text{ft}^{2}$$

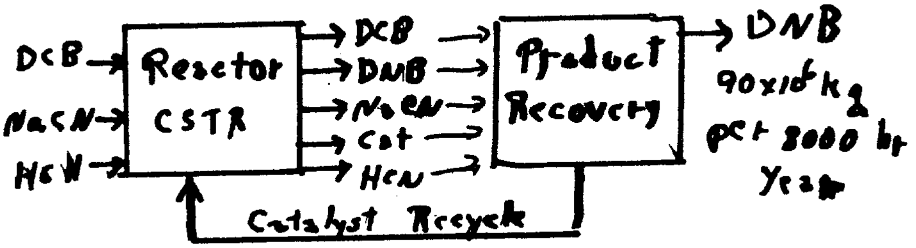
Reboiler cost $= 14{,}104 \times 25 = \underline{\underline{\overset{\$}{350{,}000}}}$

Answer

Ap C Pr: 34

Pr. Ses. Prob. 17

$$\underset{\text{M.W.}=125}{\overset{\text{DCB}}{C_4H_6Cl_2}} + 2\,\underset{\text{MW}=49}{NaCN} \xrightarrow{\text{Catalyst}} \underset{\text{M.W.}=106}{\overset{\text{DNB}}{C_4H_6(CN)_2}}$$



? = % return on extra investment for going from 80% to 85% conversion

<u>Catalyst solution</u>

	wt %
$Na_2Cu(CN)_3$ (M.W. = 138.6) =	6.5
NaCN	= 17.3
H_2O	= 76.2

Density = 1.16×10^3 kg/m^3

0.038 kg of Cu (At. wt. = 63.6) needed per kg of DCB charge

Basis: 100 kg of cat. solution = 6.5 kg $Na_2Cu(CN)_3$

$$= \frac{6.5 \mid 63.6}{\mid 138.6} = 2.98 \text{ kg Cu}$$

$$= \frac{2.98}{0.038} = 78.42 \text{ kg DCB charge/100 kg of cat. solution}$$

So need $\frac{100}{78.42} = 1.275$ kg cat. solution per kg of DCB charge

Pr. Ses. Prob. 17 (continued)

Amount of DCB feed (No side losses)

for product rate of $\frac{90{,}000{,}000}{8000} = 11{,}250$ kg DNB/hr.

amount of DCB needed in feed if yield is 80%

$$= \begin{array}{c|c|c} 11{,}250 & 125 & \\ \hline & 106 & 0.8 \end{array} = 16{,}583 \text{ kg DCB/hr}$$

amount of DCB needed in feed if yield is 85%

$$= \begin{array}{c|c|c} 11{,}250 & 125 & \\ \hline & 106 & 0.85 \end{array} = 15{,}608 \text{ kg DCB/hr}$$

Total DCB feed stream is

$$\frac{16{,}583}{1.16 \times 10^3} = 14.30 \text{ m}^3/\text{hr at } 80\% \text{ conv.}$$

$$\frac{15{,}608}{1.16 \times 10^3} = 13.46 \text{ m}^3/\text{hr at } 85\% \text{ conv.}$$

Total Catalyst stream is

$(16{,}583)(1.275) = 21{,}143$ kg cat. str./hr at 80% conv.

$$\times \frac{1}{1150} = 18.39 \text{ m}^3/\text{hr}$$

$(15{,}608)(1.275) = 19{,}900$ kg cat. str./hr at 85% conv.

$$\times \frac{1}{1150} = 17.30 \text{ m}^3/\text{hr.}$$

NaCN stream is

$$\begin{array}{c|c|c} 16{,}583 & (2)(49) & \\ \hline & 125 & 0.26 \end{array} = 50{,}004 \text{ kg/hr} \times \frac{1}{1130} = 44.25 \text{ m}^3/\text{hr}$$

for 80% conv.

$$\begin{array}{c|c|c} 15{,}608 & (2)(49) & \\ \hline & 125 & 0.26 \end{array} = 47{,}064 \text{ kg/hr} \times \frac{1}{1130} = 41.65 \text{ m}^3/\text{hr}$$

for 85% conv.

Pr. Seg. Prob. 17 (Continued) ApCPr: 36

Total feed coming in (Ignoring HCN stream)

For 80% conversion

	Vol. as m^3/hr	wt. as kg/hr	Value \$/kg	Value \$/hr
DCB Stream	14.30	16,583	0.62	10,281
Catalyst Stream	18.39	21,143	—	—
NaCN Stream	44.25	50,004	0.082	4,100
Total	76.94	87,730		\$14,381

For 85% conversion

DCB Stream	13.46	15,608	0.62	9,677
Catalyst stream	17.30	19,900	-	-
NaCN stream	41.65	47,064	0.082	3,859
Total	72.41	82,572		\$13,536

Thus: **Raw material costs per year**

For 80% conv. = 14,381 × 8000 = \$115.0×10^6

For 85% conv = 13,536 × 8000 = \$108.3×10^6

Now get cost for reactors (CSTR)

V_R = vol. of reactor, m^3

X = fractional conv. of DCB as kg DCB in feed converted per kg of DCB feed

F = feed rate of DCB as kg DCB/hr

r = kg DCB conv. in reactor/(hr)(m^3 of reactor)

Pr. Seg. Prob. 17 (Continued)

Classic C.S.T.R. Equation No. 64 (4th ed.) in Chap. 16 is

$$V_R = F\,\frac{x}{r} \quad \text{based on} \quad 0 = F - F(1-x) + (-r)V_R$$

If rate is given as $\frac{\text{kg DCB conv.}}{(\text{kg DCB feed})(\text{hr})}$, change to

$\frac{\text{kg DCB conv.}}{(\text{m}^3\text{ of reactor})(\text{hr})}$ by mult. by $\frac{\text{kg DCB feed}}{\text{m}^3\text{ of feed}}$.

$$r\,(\text{for }80\%\text{ conv.}) = \frac{0.0169 \,\big|\, 60 \,\big|\, 16{,}583}{\quad\big|\quad\big|\, 76.94} = 218.5\ \frac{\text{kg DCB}}{(\text{m}^3)(\text{hr})}$$

$$r\,(\text{for }85\%\text{ conv.}) = \frac{0.0131 \,\big|\, 60 \,\big|\, 15{,}608}{\quad\big|\quad\big|\, 72.41} = 169.4\ \frac{\text{kg DCB}}{(\text{m}^3)(\text{hr})}$$

$$V_R\,(\text{for }80\%\text{ conv.}) = \frac{16{,}583 \,\big|\, 0.8}{\quad\big|\, 218.5} = 60.7\ \text{m}^3 \times 264\ \frac{\text{gal}}{\text{m}^3} = 16{,}037\ \text{gal.}$$

$$= F\,\frac{x}{r}$$

$$V_R\,(\text{for }85\%\text{ conv.}) = \frac{15{,}608 \,\big|\, 0.85}{\quad\big|\, 169.4} = 78.3\ \text{m}^3 \times 264\ \frac{\text{gal}}{\text{m}^3} = 20{,}682\ \text{gal.}$$

Reactor cost for 80% conv. $= 260{,}000 + \frac{1.037}{9}(130{,}000)$
$= \$274{,}000 \times 4.5 = \$1{,}233{,}000.$

Reactor cost for 85% conv $= 260{,}000 + \frac{5.682}{9}(130{,}000)$
$= \$342{,}000 \times 4.5 = \$1{,}539{,}300.$

Cost for cat. solution (80% conv) $= 21{,}143 \times 2 \times 0.30 = \$12{,}685$

Cost for cat. solution (85% conv) $= 19{,}900 \times 2 \times 0.30 = \$11{,}940$

Pr Ses. Prob. 17 (Continued)

Total F.C.I. (80% Conv.) $= 1{,}233{,}000 + 12{,}685$

$\overset{\$}{=} 1{,}245{,}685$

Total F.C.I. (85% Conv.) $= 1{,}539{,}300 + 11{,}940$

$\overset{\$}{=} 1{,}551{,}240$

Total annual variable cost

For 80% conv. $= \underset{\text{Raw mtl.}}{115.0 \times 10^6} + \underset{\text{Depreciation}}{1{,}245{,}685 \times 0.08}$

$\overset{\$}{=} 115.1 \times 10^6$

for 85% conv. $= \underset{\text{\$ Raw mtl.}}{108.3 \times 10^6} + \underset{\text{Depreciation}}{1{,}551{,}240 \times 0.08}$

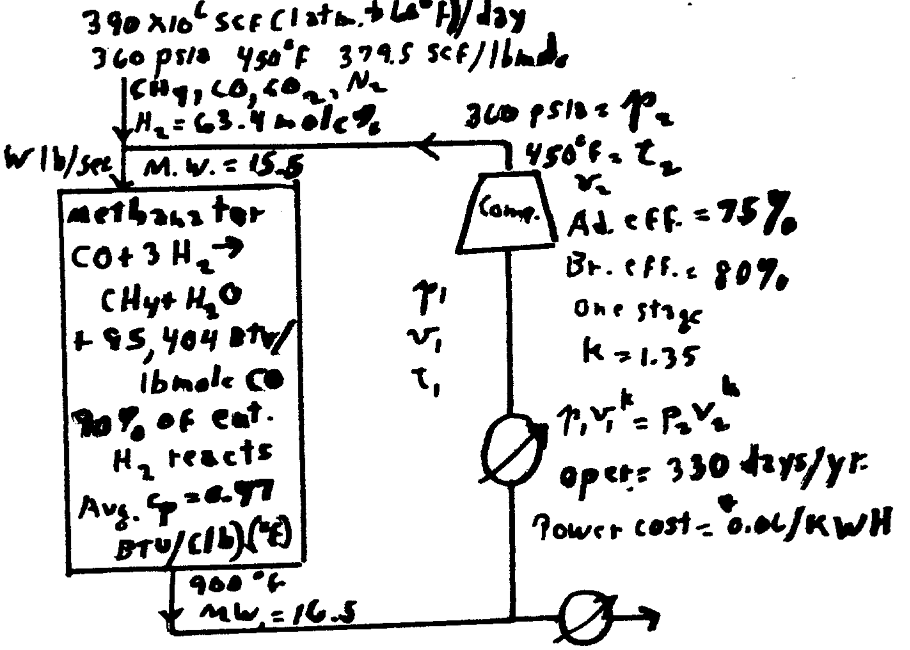
$= 108.4 \times 10^6$

% return by going from 80% to 85% Conv

$$= \frac{(115.1 - 108.4) \times 10^6}{1{,}551{,}240 - 1{,}245{,}685}(100) = \underline{\underline{22\%}}$$

Answer

Pr. Ses. Prob. 18



D of Methanator = 24.8 ft

L of Methanator = 12.4 + 3 = 15.4 ft

Theo. Hp (Eq 24-Ch. 14) (4th ed) $= \frac{3.03 \times 10^{-5} k}{k-1} p_1 q_{fm_1} \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right]$

Act. Hp. $= \frac{\text{Theo. Hp}}{0.80 \ | \ 0.75}$
Br. eff. Ad. eff.

ΔP (P. 5-53 of Perry's 6th) $= \frac{2 f G^2 L (1-\epsilon)^{3-n}}{D_p g_c \rho \phi_s^{3-n} \epsilon^3}$ lb/ft^2

$f = 1$ G = lb/(sec)(ft)^2 $L = 15.4$ ft $\epsilon = 0.4$ $n = 2$

$D_p = 0.0238$ ft $g_c = 32.17$ (lb)(ft)/(lb)(sec)^2

$\rho = 0.464$ lb/ft^3 $\phi_s = 1.0$

$i = 0.11$ $n = 20$

P.V. (Eq 24-Ch. 7) $= R \frac{(1+i)^n - 1}{i(1+i)^n}$

FCI = \$5,000,000

Pr. Sec. Prob. 18 (Continued)

?ⓐ = Power cost/yr for recycle-gas compression

?ⓑ = Net Present Value = F.C.I. + P.V. at 11% annual int. and 20-year life.

SOLUTION

Determine flow through reactor:

$$\text{Entering gas} = \frac{390\times 10^6\ \text{scf}}{24\times 3600\ \text{sec}}\left|\frac{\text{lb mole}}{379.5\ \text{SCF}}\right. = 11.9\ \frac{\text{lb mole}}{\text{sec}}$$

$11.9 \times 15.5 = 184$ lb/sec

$$\text{Lb Moles } H_2 \text{ reacting/sec} = (11.9)(0.634)(0.9)$$

$$= 6.79\ \text{lb/sec}$$

$$\text{Heat released/sec} = \frac{6.79}{\ }\left|\frac{95{,}404}{3}\right. = 215{,}936\ \frac{\text{BTU}}{\text{sec}}$$

Let W (lb/sec) = required flowrate to have temp. go from 450°F to 900°F. (No heat loss)

$$215{,}936 = W(0.47)(900-450)$$

W = 1021 lb/sec enter at 360 psia and 450°F

Flow rate through compressor = 1021 − 184 = 837 $\frac{\text{lb}}{\text{sec}}$

$$G\ (\text{for } \Delta P \text{ Eq.}) = \frac{837\ \text{lb}}{\text{sec}}\left|\frac{4}{\pi(24.8^2)\ \text{ft}^2}\right. = 1.734\ \frac{\text{lb}}{(\text{sec})(\text{ft})^2}$$

$$\Delta P = \frac{2}{0.0238}\left|\frac{1}{32.17}\right|\frac{(1.734)^2}{0.464}\left|\frac{15.4}{(1)^1}\right|\frac{(0.6)^1}{(0.4)^3} = 2444\ \frac{\text{lb}}{\text{ft}^2}$$

$$= \frac{2444}{144} = 17.0\ \text{psia}$$

$P_2 = 360$ psia $\qquad P_1 = 360 - 17 = 343$ psia

$$v_2\ (\text{at } 360\ \text{psia} + 450°\text{F}) = \frac{379.5}{\ }\left|\frac{910}{520}\right|\frac{14.7}{360}\left|\frac{\ }{16.5}\right. = 1.643\ \frac{\text{ft}^3}{\text{lb}}$$

Pr. Ses. Prob. 18 (continued)

$$v_1 = v_2 \left(\frac{P_2}{P_1}\right)^{1/k} = 1.643\left(\frac{360}{343}\right)^{\frac{1}{1.35}} = 1.703 \ \frac{ft^3}{lb}$$

$$= \frac{1.703\ ft^3}{lb}\left|\frac{837\ lb}{sec}\right|\frac{60\ sec}{min} = \underset{\overset{\|}{q_{fm}}}{85{,}522}\ \frac{ft^3}{min}$$

$$\text{Theo. Hp.} = \frac{3.03\times10^{-5}}{}\left|\frac{1.35}{0.35}\right|\frac{343\times144}{}\left|\frac{85{,}522}{}\right. \times\left[\left(\frac{360}{343}\right)^{\frac{0.35}{1.35}}-1\right] = 6228\ Hp$$

$$\text{Act. Hp} = \frac{6228}{0.80}\left|\frac{}{0.75}\right. = 10{,}380\ Hp$$

1 Hp = 0.746 kW. Plant operates 330 days/yr.

ⓐ Power cost/yr $= \dfrac{10{,}380\ Hp}{}\left|\dfrac{0.746\ kW}{Hp}\right|\dfrac{330\times24\ hr}{yr}\left|\dfrac{\$0.06}{kWhr}\right.$

$= \$3{,}680{,}000/yr$

Answer ⓐ

ⓑ Present Worth of power cost (20 years at 11%)

$$= 3{,}680{,}000\ \frac{(1+0.11)^{20}-1}{(0.11)(1+0.11)^{20}}$$

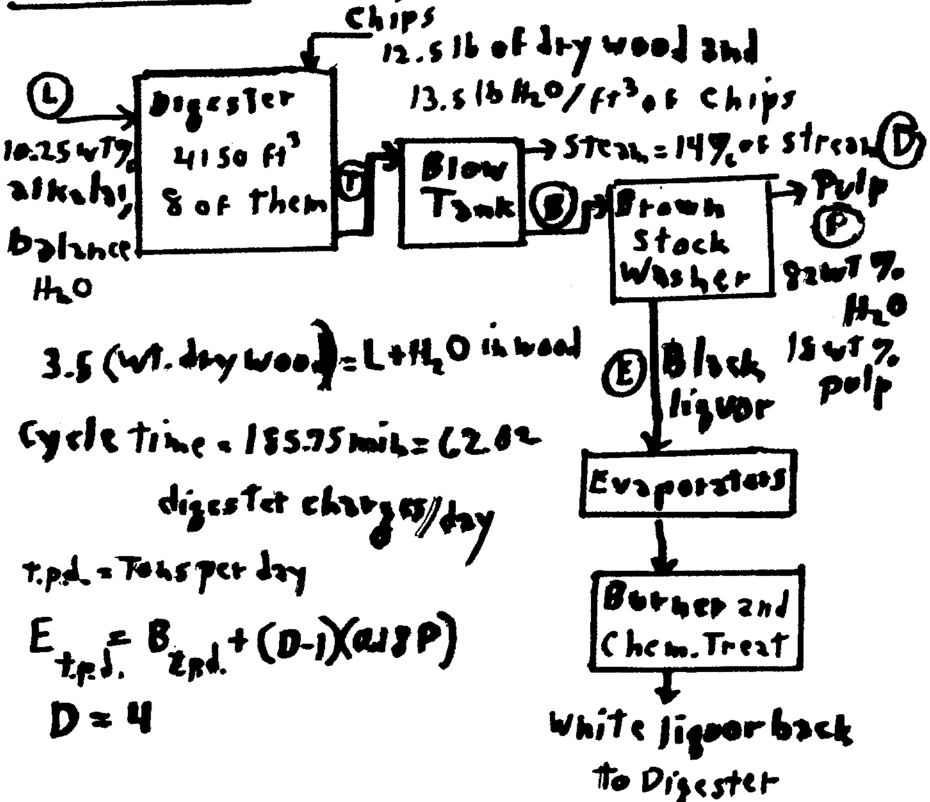
$= \$29{,}300{,}000$

Net Present Worth $= 29{,}300{,}000 + \overbrace{5{,}000{,}000}^{FCI}$

$= \$34{,}300{,}000$

Answer ⓑ

Pr. Ses. Prob 19 Ap CPr: 42



$3.5\,(\text{wt. dry wood}) = L + H_2O$ in wood

Cycle time = 185.75 min = 62.02 digester charges/day

t.p.d. = Tons per day

$E_{t.p.d.} = B_{t.p.d.} + (D-1)(0.18P)$

$D = 4$

Total charge of dry wood = $\frac{4150 \mid 12.5 \mid 62.02}{\quad\mid\quad\mid 2000}$

= 1608.6 t.p.d. dry wood

Total charge of H_2O in wood = $\frac{4150 \mid 13.5 \mid 62.02}{\quad\mid\quad\mid 2000}$

= 1737.3 t.p.d. H_2O in wood

Total wt. of wood charge = 3345.9 t.p.d. total wood

Ⓛ = wt. of liquid to Digester = 3.5 (1608.6) − 1737.3

= 3892.8 t.p.d. white liquor

Pr. Ses. Prob. 19 (Continued)

WT. of alkalai in Ⓛ = (0.1025)(3892.8) = 399 t.p.d.

WT. of H_2O in Ⓛ = (0.8975)(3892.8) = 3494 t.p.d.

WT. of stream T to Blow Tank = 3892.8 + 3345.9
= 7238.7 t.p.d

WT. of H_2O off in Blow Tank = (7238.7)(0.14) = 1013.4 t.p.d

H_2O in stream B to B.S. Washer = 1737.3 wood
+ 3494 in Ⓛ − 1013.4 in Blow
= 4217 t.p.d.

Alkalai in stream B to B.S. Washer = 399 t.p.d

Pulp in stream T and in stream B = (1608.6)(0.45)
= 724 t.p.d. unbleached pulp.

Dissolved solids wood in streams T and B =
(1608.6)(0.55) = 885 t.p.d.

Ignore the 3 t.p.d. loss of turpentine.

Total wt. of stream B = 4217.9 H_2O + 399 alk.
+ 885 dis. wood + 724 pulp = 6226 t.p.d.
(above includes 399 + 885 = 1284 t.p.d dis. solids)

Check: Ⓑ = 3345.9 + 3892.8 − 1013.4 = 6225.3 t.p.d
Check is OK

Now procede to Brown Stock Washer:

Pr. Sec. Prob. 19 (Continued)

B.S. Washer $E_{t.p.d}$ to Evap. = $B_{t.p.d}$ to B.S.W. + $(D-1)(0.18P)$

$B_{t.p.d.}$ To B.S.W. = 6226 t.p.d.

$$D = \left(\frac{\text{Ton water wash + Ton water in prod}}{\text{ton dry pulp}}\right) = 4$$

P = Ton wet pulp/day

Dry unbleached pulp = 724 t.p.d

$$P = \frac{724}{0.18} = 4022 \text{ t.p.d}$$

$E_{t.p.d}$ to Evap. = $6226 + (4-1)(724) = \underline{\underline{8398 \text{ t.pd}}}$

Answer a

WT. of dissolved solids in stream B and in stream E = 399 alk + 885 diss. wood = 1284 t.p.d

wt. % of dissolved solids in stream E = $\frac{1284}{8398}(100)$

$= \underline{\underline{15.3 \text{ wt.\%}}}$

Answer a

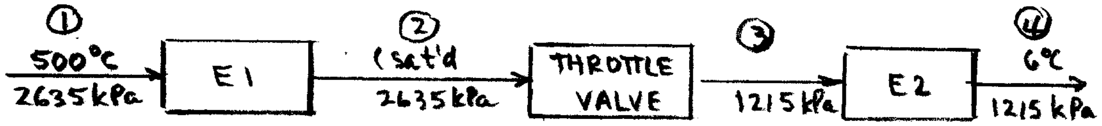
Rate of prod. of final, bleached, dry pulp

$= (0.9)(724) = \underline{\underline{652 \text{ t.p.d.}}}$

Answer b

Pr. Ses. Prob. 21

This problem is similar to one outlined by Seider, Seader and Lewin in Process Design Principles. The authors suggest using Aspen Plus to complete the material and energy balance. The flow sheet for the process is



Using the Aspen Plus subroutine HEATER to model the two heat exchangers and VALVE to simulate the action of the throttling valve the following results are obtained (these calculations use the SRK equation of state to compute the thermodynamic properties):

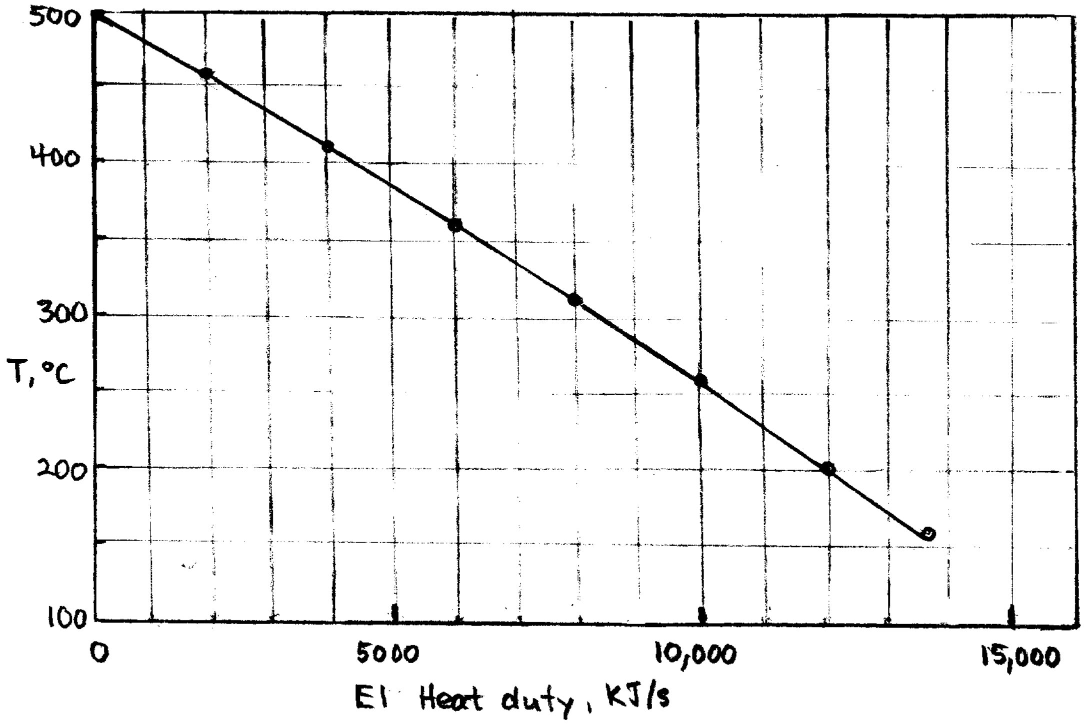
Stream	1	2	3	4
Temp., °C	500	158	140	6
Press., kPa	2635	2635	1215	1215
Mass flow, kg/s	33.37	33.37	33.37	33.37
Mol flow, kg mol/s	0.5388	0.5388	0.5388	0.5388
Enthalpy cont., kJ/s	-13,808	-27,516	-27,516	-43,045
Heat removed	E1 = 13,708 kJ/s		E_2 = 15,529 kJ/s	

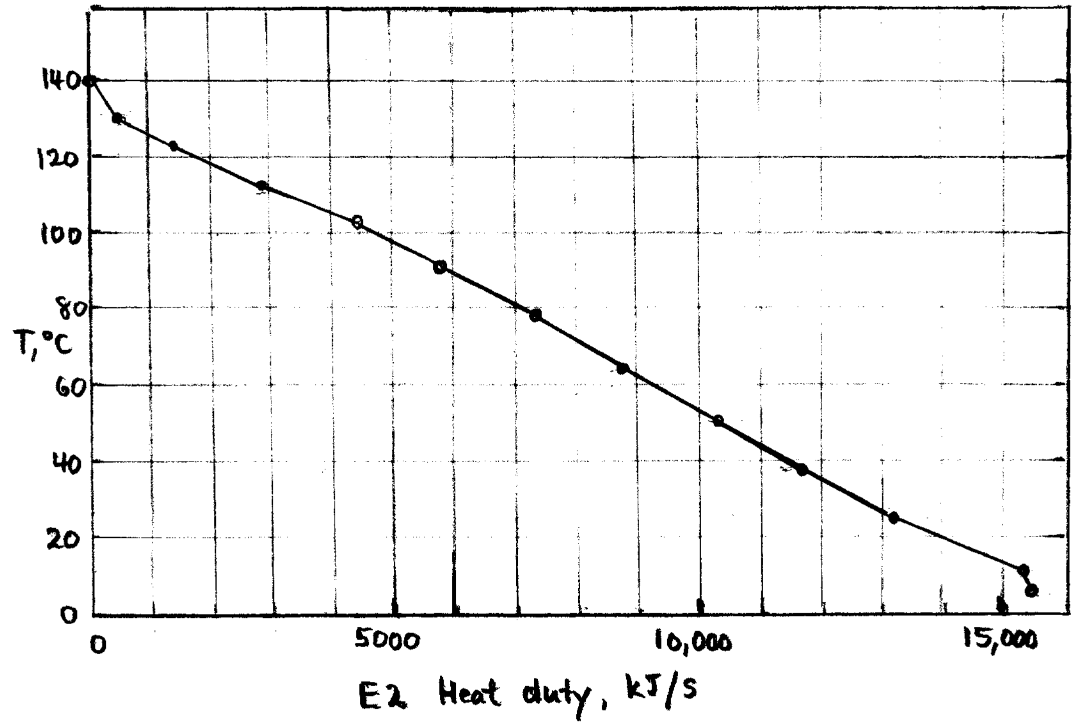
Note that the saturation temperature of the mixture at 2635 kPa is 158°C. Adiabatic expansion of this exit stream from E1 reduces the temperature to 140°C at a pressure of 1215 kPa. Thus, the exit stream upon adiabatic expansion is slightly superheated. This superheat must be removed in the entrance section of E2 as shown on the temperature diagram for this heat exchanger. Since the bubble point of the mixture is 10°C at 1215 kPa another small change occurs at this temperature to indicate subcooling of the mixture.

Ap [illegible]

Pr. Ses. Prob. 21 (cont.)

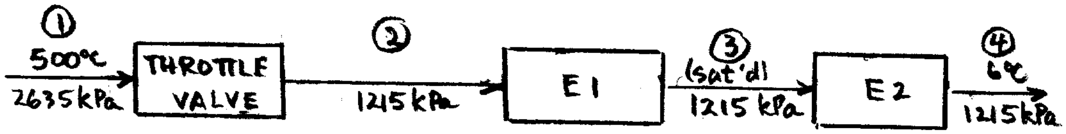
The temperature profiles in E1 and E2 are shown below:





Pr. Ses. Prob. 21 (cont.)

There are numerous arrangements that could be suggested to achieve the same product end point. One suggested arrangement below, involves reducing the pressure before entering the first exchanger. Thus,



In this alternative arrangement the inlet temperature to E1 will be more than 20°C below that for the initial arrangement. With a lower temperature driving force the area of E1 will be larger than for the initial arrangement. However, because of the lower pressure, the required thickness of the tubes in E1 may be somewhat less. The effect on E2 with this arrangement will be minimal. (Of course, E1 and E2 could be combined into one larger area exchanger.) Comparison of the two arrangements will need to be verified with an economic analysis. Without this analysis, the initial arrangement is preferred over the alternative arrangement.

Ap C Pr: 48

Pr. Session Problem 22

From an analysis of the cooling process, the hot flue gas enters at 149°C and a 37.7°C dew point (equivalent to a corresponding humidity of about 0.042 kg water/kg dry air with an enthalpy of 276 kJ/kg and an adiabatic saturation temperature of 50°C). During cooling the hot flue will proceed along the adiabatic saturation line until condensation occurs at 50°C. The cooling process then follows the saturation line down to the exit temperature of 32°C. At that point the water content is 0.031 kg water/kg dry air with an enthalpy of 116 kJ/kg. Thus, this cooling is one of dehumidification in which the mass and energy flows are in the same direction.

Calculate the total heat load

$$\dot{q}_T = \dot{m}(h_{in} - h_{out}) = (2.52\,kg/s)(276 - 116)kJ/kg$$
$$= 403\,kW$$

Latent heat total of the total heat load is the amount of water that must be condensed

$$\dot{q}_{cond} = (2.52)(0.042 - 0.031)(2402) = 66.5\,kW$$

Thus the sensible heat load is 336.5 kW or 83.5% of the total heat load.

Temperature rise of the water is 49 - 26.8 = 22.2°C, the water requirement is

$$\dot{m}_{H_2O} = 403/(4.186)(22.2) = 4.34\,kg/s$$

Assuming essentially linear temperature changes throughout the cooler, the LMTD = 32°C.

For this system the liquid-to-gas flow rate ratio is

4.34/2.52 = 1.72

For flow ratios of 1.5 to 3 in direct gas-liquid heat exchange, a packed tower is generally selected to effect the heat transfer and maintain a reasonable pressure drop. See Table 3-18 of D.R. Woods, *Process Design and Engineering Practice*, p. 3-73, Prentice Hall (1995).

To size the tower assume a gas velocity of 1 m/s. Since the density of the flue gas changes from about 0.8 kg/m^3 at the inlet to about 1.1 kg/m^3 at the exit, assume an average of about 1 kg/m^3.

Cross-sectional area required with these assumptions is

$$A = \dot{m}/\rho V = 2.52/(1)(1) = 2.52\,m^2$$

Pr. Ses. Prob. 22 (cont.)

or

$$D = (A/0.785)^{1/2} = (2.52/0.785)^{1/2} = \underline{1.79\,m}$$

ANSWER

The gas and liquid loadings for the cooler are:

$$\text{gas}: 2.52\,kg/s\ /\ 2.52\,m^2 = 1\,kg/s\cdot m^2$$

$$\text{liquid}: 4.34/2.52 = 1.72\,kg/s\cdot m^2$$

The overall volumetric heat transfer coefficient is obtained by combining a gas and liquid coefficient. Woods provides a value of 3 kW/m^2•°C for a typical volumetric overall heat transfer coefficient for a packed tower under the above conditions (see Table 3-18). For greater accuracy values for the individual gas and liquid volumetric heat transfer coefficient may be obtained using the relations

$$h_V = 3(\dot{m}_V/A)(\dot{m}_L/A)^{0.2}$$

$$h_L = 42(\dot{m}_V/A)^{0.7}(\dot{m}_L/A)^{0.5}$$

These values can be used to obtain the overall volumetric heat transfer coefficient.

Using the value of 3 kW/m^2•°C for U, the volume can be determined from

$$\text{volume} = \dot{q}_T / \upsilon\Delta T_{\ln m}$$

$$= (403)/(3)(32) = 4.2\,m^3$$

$$L = 4.2/2.52 = \underline{1.67\,m}$$

ANSWER

This provides essentially a 1:1 ratio for the L/D ratio.

The pressure drop in the cooler is obtained directly from using Fig. 2-17 provided by Woods. For Pall rings this would be approximately 0.35kPa/m. Thus, the total pressure drop in the cooler with a length of 1.67 + 0.2 = 1.87 m. (the 0.2 m is added to provide sufficient clearance in the packed tower) is

$$\Delta p = (1.87)(0.35) = 0.65\,kPa$$

This is within the 5 kPa pressure drop allowed.

Pr. Ses. Prob. 22 (cont.) ApC Pr: 50

The approximate cost of a carbon steel tower with these dimensions (requires an extrapolation of the cost curve of Fig. 15-11) without any packing is $12,000 on Jan. 2002.

ANSWER

Note: This problem was originally developed by H. Sawitstowski and W. Smith, *Mass Transfer Process Calculations*, p. 456, Interscience Publishers, New York (1963). In these calculations, they arrived at a total heat load of 405 kW, a calculated volumetric heat transfer coefficient of 2.4 $kW/m^2 \cdot {}^\circ C$, a diameter of 1.83 m and a height of 1.98 m.

Pr. Session Problem 23

ApC Pr :51

To provide a basis for devising alternative schemes for creating the streams required, list the components in increasing order of boiling temperature and note the various characteristics of the components. Such an analysis shows a very wide range in boiling temperatures, from −196 to 98°C. All components have similar solubility parameters, and are non-polar. Hydrogen sulfide is the only component that is fairly corrosive. Except for CO_2 and N_2 most of the components have similar flammability ratings.

To obtain the specific streams required, use a heuristic approach to suggest possible separation schemes. Note that some of the heuristics will provide conflicts that will require further analysis before a decision on the separation scheme is proposed. (See Appendix F).

1. Remove the most hazardous, corrosive or unstable component first. In this case this would be hydrogen sulfide.
2. Save the most difficult separation until last. The easiest to separate are those that exhibit the greatest difference in molar mass or between boiling temperatures, or between freezing temperatures. Several of the molar masses are within the 30 to 45 range. The largest difference in temperature occurs between the N_2 and C_1 and the rest of the components.
3. Separate the most plentiful components early in the process. Here CO_2 should be removed early because it sublimes that can result in blocked lines and because if forms an azeotrope with C_2.
4. Try to avoid adding an agent to achieve the separation. There appears to be a reasonable difference in boiling temperature except for the most plentiful substance, CO_2, which sublimes. According to heuristic#3, CO_2 should be separated early. Thus, distillation might be useful for all separations except the separation of CO_2.
5. Leave high specific recoveries until last.

6. Avoid extremes in operating conditions. Atmospheric pressure operation is acceptable for iC_3 and higher. For the other species we should require a higher pressure and this is available with the inlet pressure of 2 MPa.

7. When considering separation options, recognize safety, controllability, and materials of construction integration as well as energy conservation.

Note that some of the heuristics offer some conflicts. For example, heuristic#1 recommends removal of the H_2S first; heuristic#3 says separate the CO_2 early; heuristic#2 says separate C_1 from C_2 early, or separate CO_2 from C_2 last. However, removing the corrosive component first makes the most sense. This will probably involve absorbing H_2S preferentially in an absorbing liquid. Such H_2S removal is used in every one of the four proposed options available to the design engineer. In each separation option, assume the H_2S is removed first.

Option 1:

Separate the C_1 cut first: use distillation without reflux and without the temperatures that might cause CO_2 freezing. Do not use a solvent in the first column. This allows more CO_2 to go overhead and the latter is lost by scrubbing the methane to free it from the CO_2. The bottoms is sent to an extractive distillation column. Figure 23a shows the variation. This option is shown in Fig. P-23a.

Option 2:

Concentrate the CO_2 using membrane technology. Distill the retentate to remove the methane. Take the bottoms from the demethanizer and perform an azeotropic distillation to separate CO_2 from C_2 using a C_4 extraction fluid. Treat the azeotrope overhead with a membrane distillation hybrid to remove the CO_2. Now combine the various streams as shown in Fig. P-23b.

Option 3:

Separate the C_1 cut first using simple distillation with some C_4 solvent added near the top of the distillation column to prevent the CO_2 from freezing. This separates the C_1 from the rest of the feed. The bottoms goes to an extractive distillation column with the C_4 must used as the solvent to separate the CO_2 from the C_2. This option is shown in Fig. D-23c.

Option 4:

Partially separate the C_2 / CO_2 with an absorption process using a C_4 stream as the absorbent. With the overhead, a CO_2 / C_1 separation is pursued. The bottoms stream is separated in an absorption column using a C_4 stream to serve as the absorption stream. This configuration is outlined in Fig. P-23d.

Additional information on process design configurations is given by D.R. Woods, *Process Design and Engineering Practice: Selecting and Sizing Homogenous Separations*, McMaster University, Hamilton, ON, Canada (1993).

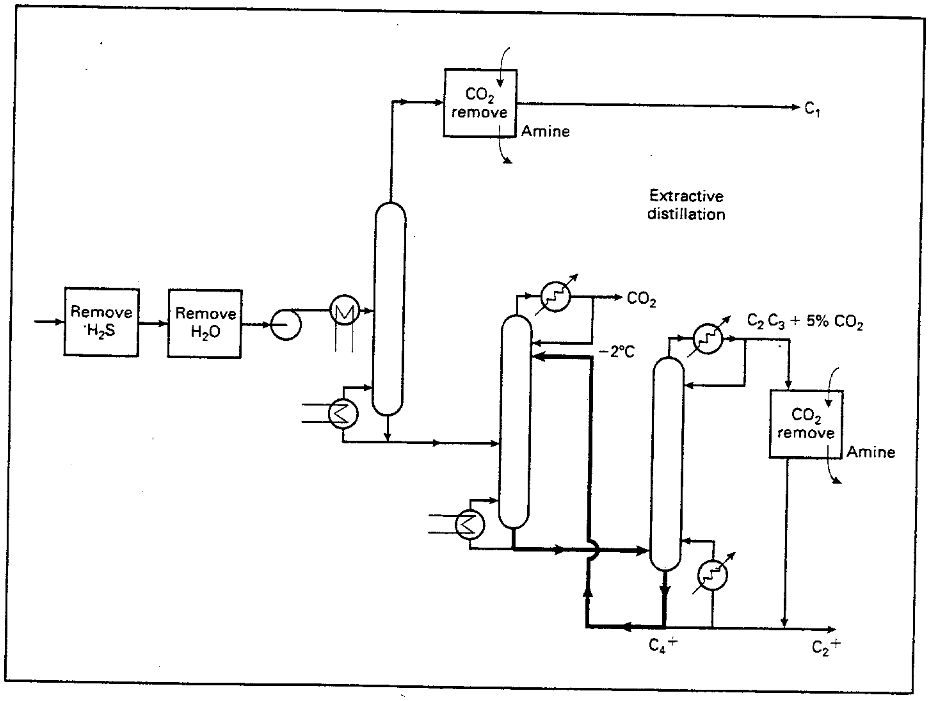


Figure P23a.

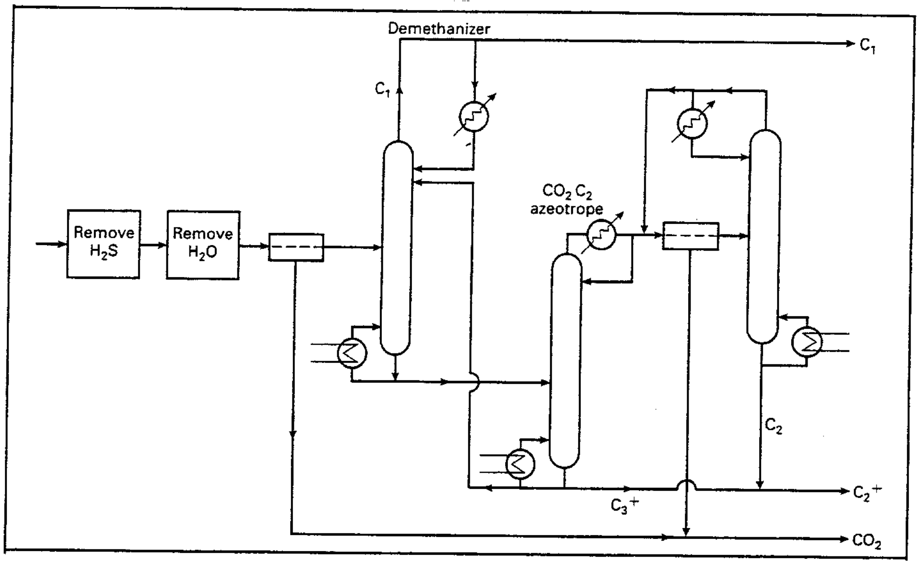


Figure P23b.

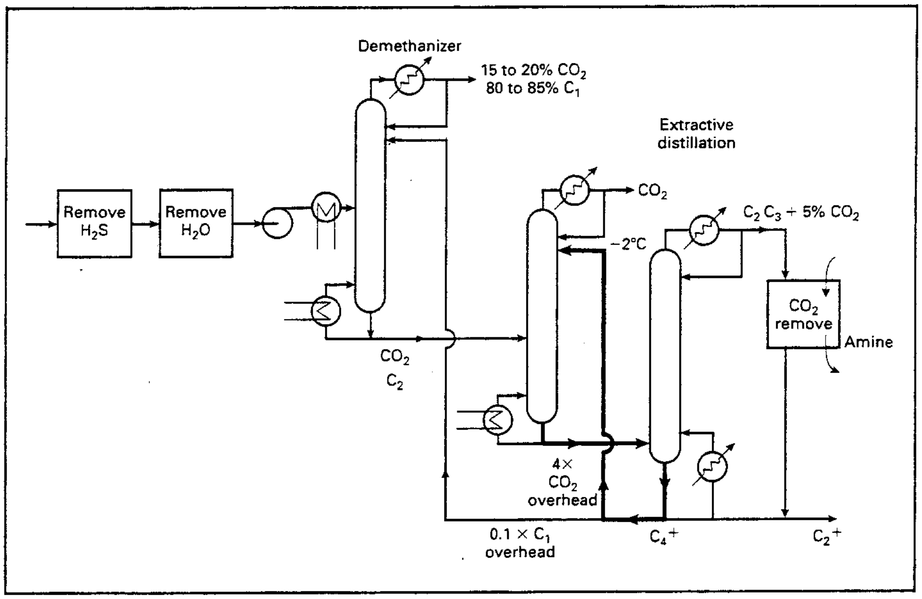


Figure P23c.

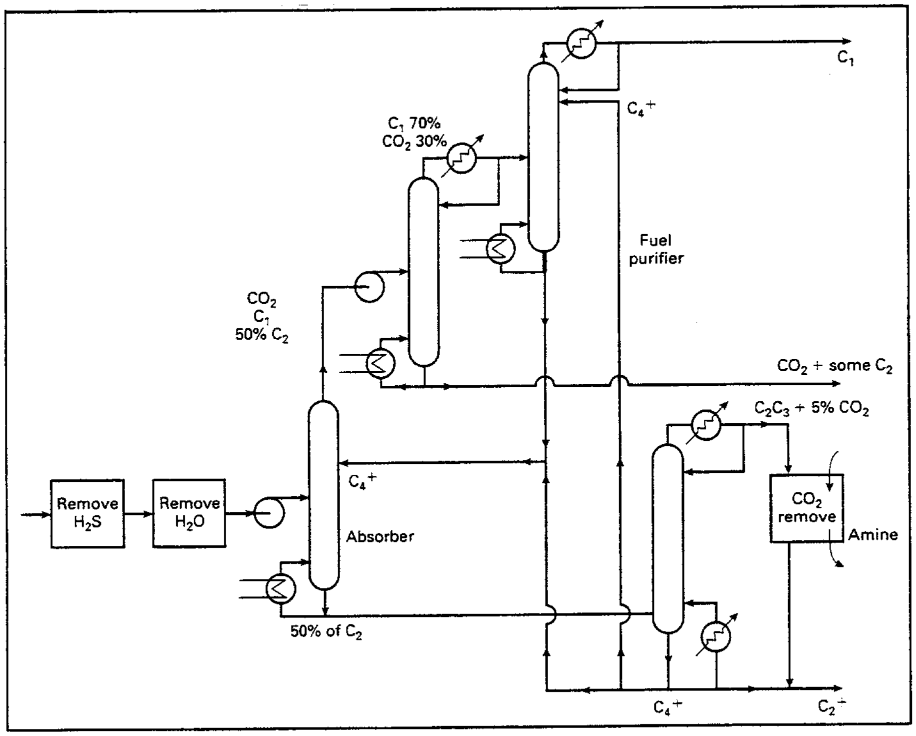


Figure P23d.

ApC Pr:56

Pr. Session Problem 24

(a) For the base case, the superficial permeate velocity is

$$\mathrm{v} = \frac{40L}{\cancel{m}^2\cancel{h}}\ \frac{\cancel{h}}{3600\,s}\ \frac{10^{-3}m^3}{L} = 1.11\times10^{-5}m/s$$

Assume essentially a water solution with a kinematic viscosity of $1\times10^{-6}m/s$, $D_i = 5\times10^{-11}m^2/s$

$$\mathrm{Sc} = \frac{\mu}{\rho D_i} = \frac{1\times10^{-6}}{5\times10^{-11}} = 20,000$$

From T.K. Sherwood, R.L. Pigford, and C.R. Wilkie, *Mass Transfer*, p. 169, McGraw-Hill, New York (1975) a relationship between Sc, Re, and Sh is given by

$$\begin{aligned}\mathrm{Sh} &= 0.096\,\mathrm{Re}^{0.913}\,\mathrm{Sc}^{0.346}\\ &= (0.0096)(25,000)^{0.913}(20,000)^{0.346}\\ &= 3060\end{aligned}$$

The mass transfer coefficient k_c based on concentration is given by

$$\mathrm{k_c} = \frac{\mathrm{Sh}D_i}{\mathrm{D}} = \frac{(3060)(5\times10^{-11})}{0.015} = 1.02\times10^{-5}m/s$$

The rejection rate R in ultrafiltration is related to the equivalent partition function K by the relation

$$\frac{1-R}{R} = \left(\frac{K}{1-K}\right)\exp\left(\frac{v}{k_c}\right)$$

Foro this relation as v/k_c approaches zero, the rejection approaches 1-K and is designated as R_{max}. Substituting in the previous equation

$$\frac{1-0.75}{0.75} = \frac{\mathrm{K}}{1-\mathrm{K}}\exp\frac{1.11\times10^{-5}}{1.02\times10^{-5}}$$

$$\mathrm{K} = 0.112/1.112 = 0.101$$

If the flux is reduced to $20\text{L}/m^2 \cdot h$ (half of the original flux)

$$v = 1.11\times10^{-5}/2 = 0.555\times10^{-5}\, m/s$$

$$\frac{1-R}{R} = \frac{0.101}{1-0.101}\exp\frac{0.555\times10^{-5}}{1.02\times10^{-5}}$$

Solving for R gives

$$R = 0.84$$

R_{macx} is obtained by allowing the flux to decrease to zero. Then

$$R_{max} = 1 - K = 1 - 0.101 = \underline{0.90}$$

Answer (a)

(b) The diffusivity of large molecules varies with the −1/3 power of the size and the mass transfer coefficient k varies with $D_i^{-0.65}$ or $M^{-0.22}$.

Predict a new partition function K_2 from K_1 where

$$K_1 = 0.101 = (1-\lambda_1)^2$$

where λ is the ratio of molecular size to pore size. For the case in part (a), $\lambda_1 = 0.682 = D_1/D_{pore}$

$$D_2 \cong D_1\left(\frac{10,000}{30,000}\right)^{1/3} = 0.693\,D_1$$

$$\lambda_2 = (\lambda_1/D_1)D_2 \qquad \lambda_2 = (0.682)(0.693) = 0.473$$

$$K_2 = (1-0.473)^2 = 0.278$$

From above

$$k_{c,2} = k_{c,1}\left(\frac{10,000}{30,000}\right)^{-0.22} = (1.02\times10^{-5})(1.27)$$

$$= 1.30\times10^{-5}\, m/s$$

At a $v = 1.11\times10^{-5}\, m/s$

Pr. Ses. Prob. 24 (cont.)

$$\frac{1-R_2}{R_2}=\left(\frac{0.278}{1-0.278}\right)\exp\frac{1.11\times10^{-5}}{1.30\times10^{-5}}=0.904$$

$$R_2 \cong \underline{0.525}$$

Answer

(c) Often the diffusion effect can be considerable if the selective layer thickness is very thin, i.e., between 0.1 to 0.2 μm as in this case. The basic equation for solute flux in the selective layer is similar to that for mass transfer in the boundary layer, but the diffusion term adds to the convective term.

$$vc_2 = vc - D_e\frac{dc}{dy}$$

where the effective diffusivity D_e is given by $D_{pore}\varepsilon/\tau$. Here D_{pore} is the diffusivity in the pores, ε the porosity fraction, and τ the tortuosity factor. Assuming the same parttion function at both boundaries of the selective layer, the differential equation can be integrated to

$$\frac{c_2}{c_s}=\frac{K\exp(vL/D_e)}{K-1+\exp(vL/D_e)} \qquad \text{L=thickness of the selective layer}$$

For a $D_{pore}=1\times10^{-11}m^2/s,\ \varepsilon=0.5,\ \tau=2,\ M=30{,}000, \text{and}\ D_i=5\times10^{-11}m^2/s$

$$D_e=(1\times10^{-11})(0.5)/2=2.5\times10^{-12}m^2/s$$

$$\frac{vL}{D_e}=\frac{(0555\times10^{-5})(2\times10^{-7})}{2.5\times10^{-12}}=0.444$$

Thus $\dfrac{c_2}{c_3}=\dfrac{(0.101)\exp(0.444)}{(0.101-1+\exp(0.444)}=0.238$

Molecular diffusion has a significant effect on the rejection since without it

$$\frac{c_2}{c_3}=0.101$$

Thus, the permeate concentration would be more than twice as high, indicating that the partition coefficient would be lower than that estimated in part (a).

Pr. Ses. Prob. 25

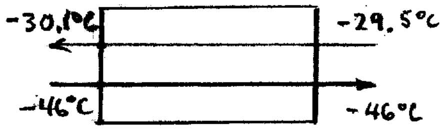
tion requirement. E603 and E604, as well as E608 and E609 can be eliminated because of the large heat exchange loads that could not be provided by the remaining heat exchangers. For example, E612 could provide less than 10% of the heat load required in E609 and necessitate two additional heat exchangers. The heal load available in E612 would most likely be better utilized in another section of the ethylene plant. E610 can also be eliminated because of the requirement of a cooling medium below 35°C.

This elimination of the above heat exchangers from those shown on Fig. 3-13 leaves exchangers E605, E606, E607, and E611 to be used in the heat integration. Using the procedure described in Chap. 9, a pinch technology plot can be developed as shown on the attached plot. This graphical summary of enthalpy rate change as a function of temperature indicates that the stream arrangement for E605 is adequate and should not be changed. The heat load from E606 can be partially used to heat up the stream in E611. This will reduce the cold water requirement in E606 and the low pressure steam requirement in E611. It will necessitate the addition of two heat exchangers to take care of the additional heat loads specified. There are no additional heat exchange opportunities to accept the heat load of E607. Just as with E612, part of the heat load could have been provided to E609. It would be better to find a more suitable heat exchange somewhere else.

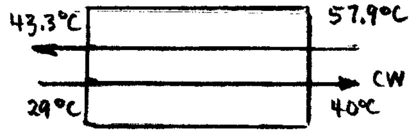
Pr. Ses. Prob. 25 (cont.)

The conditions for the 12 heat exchangers listed in Fig. 3-13 are listed below:

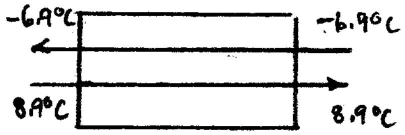
E601



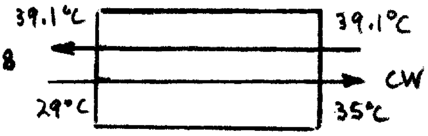
E607



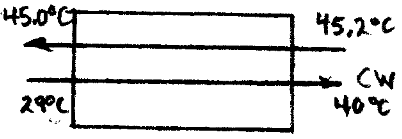
E602



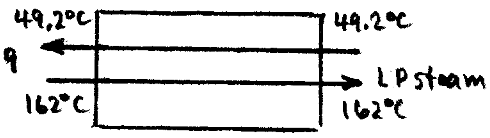
E608



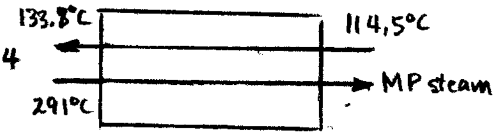
E603



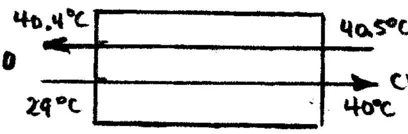
E609



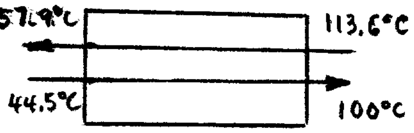
E604



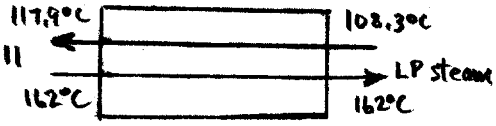
E610



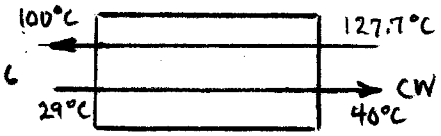
E605



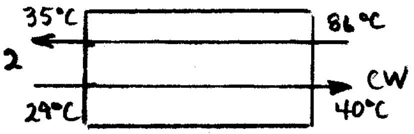
E611



E606



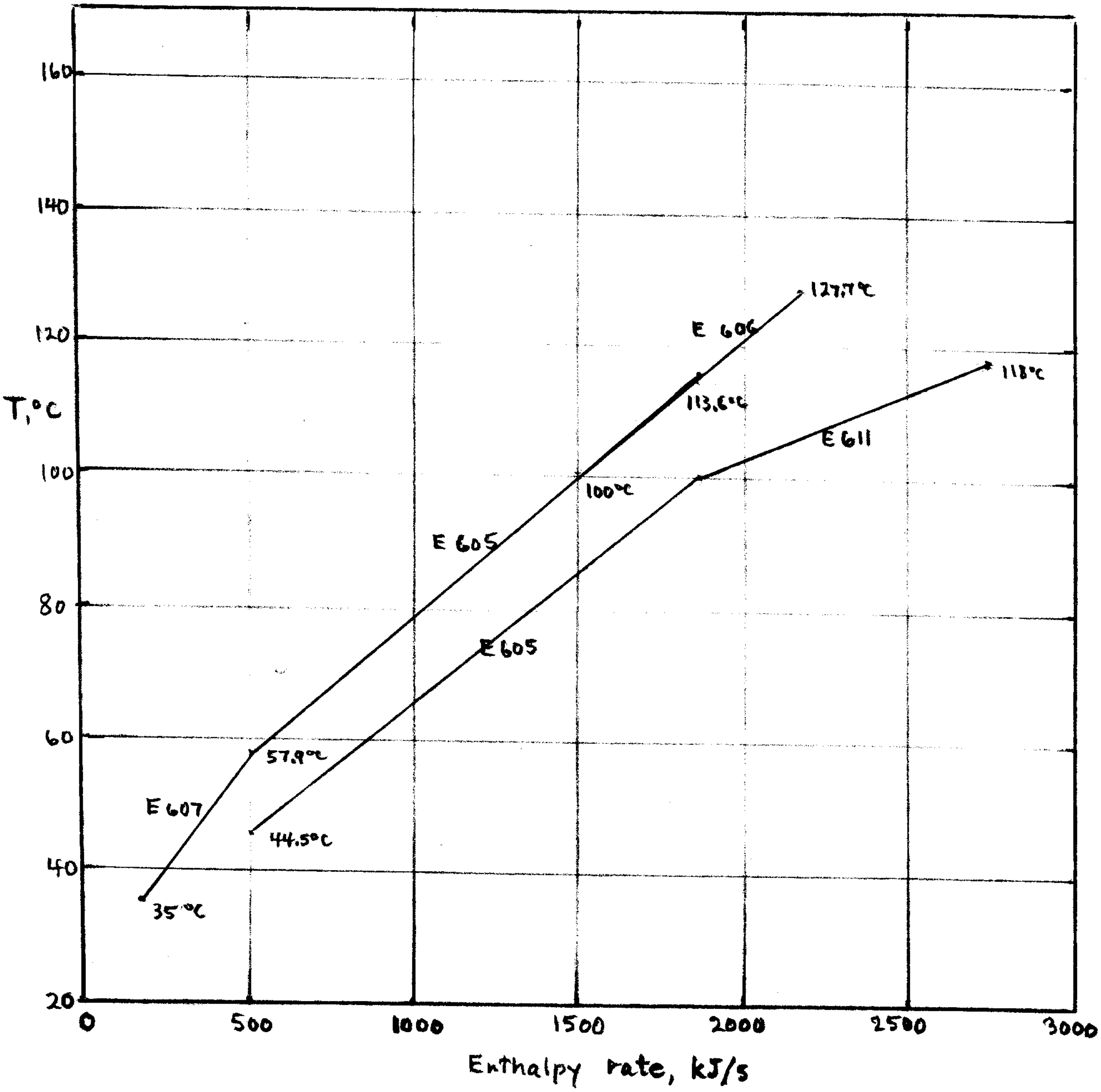
E612



A quick review of the temperature and load requirements for some of the heat exchangers makes it possible to eliminate several of the heat exchangers for the heat integration. E601 and E602 can be eliminated because of the C_3 refrigera-

Pr. Ses. Prob. 25 (cont.)

ApC Pr: 61



Pr. Ses. Prob. 26 ApC Pr: 62

Recognize that NPW can be expressed in terms of y_2, the $GeCl_4$ mol fraction in the scrubber effluent as,

$$NPW = (\$100/kg)(0.633\ kg\ mol/min.)(214.39\ kg/kg\ mol)(60\times24\times335\ operating\ min/y)(y_{2,present} - y_{2,new})(1-0.35)$$

$$-\ 1.41\left[(18{,}590) + (7{,}160)(0.40)\left(\ln\frac{0.001475}{y_{2,new}}\right)\right]$$

To maximize NPW w.r.t. $y_{2,new}$, differentiate and set $=0$. Recognize that although $y_{2,present}$ is not stated, it is a constant, and therefore drops out upon differentiation.

$$\frac{d(NPW)}{d(y_{2,new})} = (\$4.25\times10^{10})\left(\overset{0}{\cancel{\frac{d(y_{2,present})}{d(y_{2,new})}}} - \overset{-1}{\cancel{\frac{d(y_{2,new})}{d(y_{2,new})}}}\right)$$

$$-\ 4038\frac{d}{dy_{2,new}}\left(\overset{0}{\cancel{\ln(0.001475)}} - \overset{-1/y_{2,new}}{\cancel{\ln y_{2,new}}}\right)$$

$$0 = -4.25\times10^{10} + 4038/(y_{2,new})$$

$$y_{2,new} = \underline{\underline{9.5\times10^{-8}}}$$

ANS.

Note, this gives $z = 3.86\ m$