

This document contains the tutorial questions for the 15-credit Offshore Process Engineering course delivered at the University of Aberdeen.

The recommended questions you should solve for each week are:

- Tutorial 1 Phase Behaviour: Q. 1 to Q. 7.
- Tutorial 2 Single Phase Fluid Flow: Q. 10 to Q. 20.
- Tutorial 3 Separation: Q. 23
- Tutorial 4 Produced Water: Q. 24 to Q. 26.
- Tutorial 5 Compression: Q. 27 to Q. 29.
- Tutorial 6 Gas Treatment: Q. 32 to Q. 35.
- Tutorial 7 Multi-phase: Q. 39 to Q. 43.
- Tutorial 8 Hydrates: Q. 45 to Q. 47.
- Tutorial 9 Energy Efficiency: Q. 48 to Q. 57.
- Tutorial 10 Water Injection: Q. 60 to Q. 61.

All other questions are provided for additional practice (and many are past exam questions) and should help you to explore all aspects of the course.

Fully worked solutions are available but you should attempt the problems without the solutions, it's the only way to find out what you don't know!

Where marks are given, these are indicative of the *relative* weighting each part of a question might have. Please note, the number of questions in an exam (and exam durations) have changed over the years, so the overall marks for a question may now be different to what is reported here.

When answering questions, state any assumptions and reference any information sourced from literature.

1 1: Phase behavior and physical properties

Q.1

Question 1

Calculate the molar volume of carbon dioxide at 400 K and 50 bara:

- using the Peng-Robinson equation of state [$598 \text{ cm}^3 \text{ mol}^{-1}$]
- assuming the ideal gas law [$665 \text{ cm}^3 \text{ mol}^{-1}$]

Repeat the calculations at 293 K and 1 bara. Source the required value from literature and compare with your calculation.

[$24215 \text{ cm}^3 \text{ mol}^{-1}$, $24361 \text{ cm}^3 \text{ mol}^{-1}$]

Solution:

Required Data:

- $T_C = 304.2$ K (Perry's Chem Eng Handbook)
- $P_C = 73.83$ bara (Perry's Chem Eng Handbook)
- $R = 83.145 \text{ cm}^3 \text{ bar mol}^{-1} \text{ K}^{-1}$
- $\omega = 0.224$ (acentric factor from slides, or other as long as it is referenced)

The Peng-Robinson equation of state is

$$P = \frac{RT}{V - b} - \frac{a\alpha}{V^2 + 2bV - b^2} \quad (1)$$

and the polynomial compressibility form is given as

$$z^3 - (1 - B)z^2 + (A - 3B^2 - 2B)z - (AB - B^2 - B^3) = 0 \quad (2)$$

where z is the compressibility. The required parameters are given by the following relationships:

$$a = 0.45724 \left(\frac{R^2 T_C^2}{P_C} \right) \quad (3)$$

$$b = 0.07780 \left(\frac{R T_C}{P_C} \right) \quad (4)$$

$$\alpha = (1 + (0.37464 + 1.54226\omega - 0.26992\omega^2)(1 - T_r^{0.5}))^2 \quad (5)$$

$$A = \frac{a\alpha P}{R^2 T^2} = 0.45724 \left(\frac{\alpha P_r}{T_r^2} \right) \quad (6)$$

$$B = \frac{b P}{R T} = 0.07780 \left(\frac{P_r}{T_r} \right) \quad (7)$$

where subscript r denotes a reduced property.

Peng-Robinson equation of state

First, a and b are calculated:

$$a = 0.45724 \left(\frac{83.145^2 \cdot 304.2^2}{73.83} \right) = 3961886.615 \quad (8)$$

$$b = 0.07780 \left(\frac{83.145 \cdot 304.2}{73.83} \right) = 26.65275 \quad (9)$$

The reduced temperature is $T_r = T/T_C = 400/304.2 = 1.3149$ which is used to calculate α :

$$\alpha = (1 + (0.37464 + 1.54226 \cdot 0.224 - 0.26992 \cdot 0.224^2)(1 - 1.3149^{0.5}))^2 = 0.80341 \quad (10)$$

The molar mass V is now the only remaining unknown in Eq. 1. We can therefore use a calculator or excel (goal seek) etc. to solve for $V = 598 \text{ cm}^3 \text{ mol}^{-1}$.

Ideal gas law

The ideal gas law is

$$PV = RT \quad (11)$$

We can calculate the molar volume by simply plugging in the values that we are given:

$$V = \frac{RT}{P} = \frac{83.145 \cdot 400}{50} = 665 \text{ cm}^3 \text{ mol}^{-1} \quad (12)$$

There is an 11 % difference between P-R and the ideal gas law.

293 K and 1 bara

Repeat the steps above for the new temperature and pressure values.

$$T_r = \frac{293}{304.2} = 0.963182 \quad (13)$$

and

$$\alpha = (1 + (0.37464 + 1.54226 \cdot 0.224 - 0.26992 \cdot 0.224^2) (1 - 0.963182^{0.5}))^2 = 1.026434 \quad (14)$$

where a and b remain unchanged. We now solve again and obtain $V = 24215 \text{ cm}^3 \text{ mol}^{-1}$ using P-R. Using the ideal gas law, $V = RT/P = 83.145 \cdot 293/1 = 24361 \text{ cm}^3 \text{ mol}^{-1}$. As to be expected, there is close agreement between P-R and the ideal gas law at standard conditions.

[Question end]

Q.2**Question 2**

Calculate the saturation pressure for ethane at 293 K using the SRK equation of state. Compare with published information. [38 bara]

Solution:

Required Data:

- $T_C = 305.6 \text{ K}$
- $P_C = 48.3 \text{ bara}$
- $R = 83.145 \text{ cm}^3 \text{ bar mol}^{-1} \text{ K}^{-1}$
- $\omega = 0.1064$ (acentric factor)

The Soave-Redlich-Kwong equation of state is:

$$P = \frac{RT}{V - b} - \frac{a\alpha}{V(V + b)} \quad (15)$$

with polynomial compressibility form

$$z^3 - z^2 + (A - B - B^2)z - AB = 0 \quad (16)$$

The required parameters are given by the following relationships:

$$a = 0.42748 \left(\frac{R^2 T_C^2}{P_C} \right) \quad (17)$$

$$b = 0.08664 \left(\frac{RT_C}{P_C} \right) \quad (18)$$

$$\sqrt{\alpha} = 1 + m (1 - T_C^{0.5}) \text{ with } m = 0.48 + 1.57\omega - 0.172\omega^2 \quad (19)$$

$$A = \frac{a\alpha P}{R^2 T^2} = 0.42748 \left(\frac{\alpha P_r}{T_r^2} \right) \quad (20)$$

$$B = \frac{bP}{RT} = 0.08664 \left(\frac{P_r}{T_r} \right) \quad (21)$$

Calculate the reduced temperature,

$$T_r = \frac{T}{T_c} = \frac{293}{305.6} = 0.9588 \quad (22)$$

which we require in order to calculate α

$$\alpha = (1 + (0.48 + 1.574 \cdot 0.1064 - 0.172 \cdot 0.1064^2) (1 - 0.9588^{0.5}))^2 = 1.027 \quad (23)$$

Calculate A and B :

$$A = 0.42748 \left(\frac{1.027P}{0.95877^2 P_c} \right) = 0.009889P \quad (24)$$

$$B = 0.08664 \left(\frac{P}{0.95887 P_c} \right) = 0.001871P \quad (25)$$

Now substitute the expressions for A and B into the polynomial compressibility form of the SWK equation of state. This is a cubic equation, therefore, for a given value of P , there will be three compressibility roots which solve the equation. In this case the first root corresponds to the liquid phase and the third root corresponds to the vapour phase.

We can calculate the fugacity coefficient of the vapour and liquid phase using the following expression

$$\ln \phi = z - 1 - \ln(z - B) - \frac{A}{B} \ln \left(1 + \frac{B}{z} \right) \quad (26)$$

At equilibrium, the fugacities of the two phases will be equal. Solve for the pressure at which the fugacities are equal. The result will be $P_V \approx 37.3$.

[Question end]

Q.3

Question 3

Given the following hydrocarbon composition calculate the dew and bubble point curves from 1 bara to 100 bara. Use the Wilson expression to generate your K values.

$$K_i = \frac{P_{ci}}{P} e^{[5.37(1+\omega_i)(1-T_{ci}/T)]} \quad (27)$$

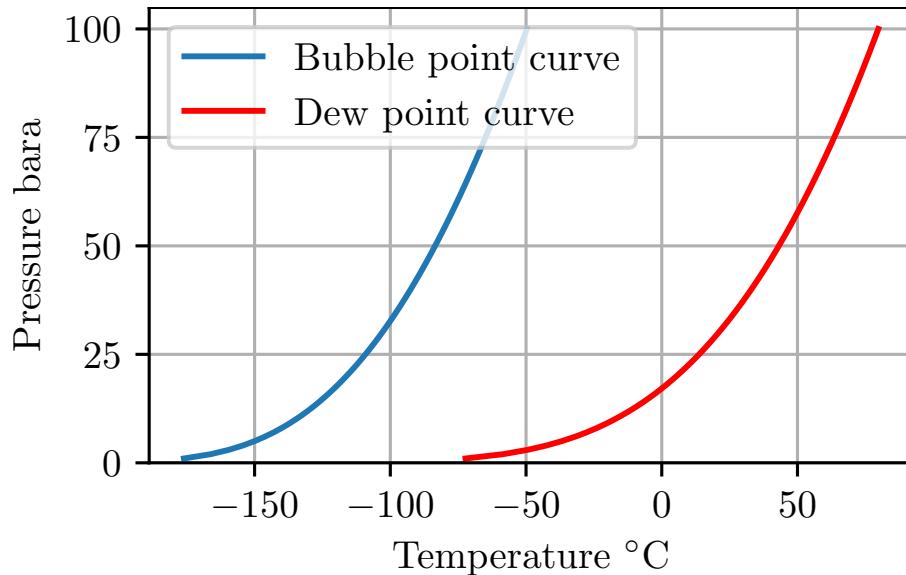
where P is the absolute pressure, P_{ci} is the critical pressure of component i in the same units as P , T is the absolute temperature, T_{ci} is the critical temperature of component i in the same units as T , and ω_i is the acentric factor of component i .

| Component | Mol fraction |
|-----------|--------------|
| Methane | 0.65 |
| Ethane | 0.15 |
| Propane | 0.1 |
| i-Butane | 0.05 |
| n-Butane | 0.05 |

Use the following properties:

| Component | P_c bara | T_c °C | T_c K | Acentric factor |
|-----------|------------|----------|---------|-----------------|
| Methane | 46.39 | -82.45 | 190.55 | 0.0115 |
| Ethane | 48.83 | 32.28 | 305.28 | 0.0986 |
| Propane | 42.55 | 96.75 | 369.75 | 0.1524 |
| i-Butane | 36.46 | 134.9 | 407.9 | 0.1848 |
| n-Butane | 37.95 | 152 | 425 | 0.201 |

Answer:



Solution:

Dew point curve

At the dew point the mixture needs to satisfy

$$\sum_i \left(\frac{y_i}{K_i} \right) = 1.0 \quad (28)$$

Now for a given P , find the temperature which satisfies the above equation.

For example, for a pressure $P = 40.7$ bar and temperature $T = 173$ °C we get the following values:

$$K_{methane} = 0.6569 \quad (29)$$

$$K_{ethane} = 0.01318 \quad (30)$$

$$K_{propane} = 0.0009178 \quad (31)$$

$$K_{i-butane} = 0.0001586 \quad (32)$$

$$K_{n-butane} = 7.756 \times 10^{-5} \quad (33)$$

Finally:

$$\sum_i \left(\frac{y_i}{K_i} \right) = 1081.13 \quad (34)$$

Solving for the value which satisfies Eq. 28 gives a temperature $T \approx 327$ K. Repeat this calculation for a range of pressure values in order to obtain the dew point curve.

Bubble point curve

At the bubble point the mixture must satisfy

$$\sum_i K_i x_i = \sum_i y_i = 1.0 \quad (35)$$

We must repeat the process above and find a temperature T which satisfies this new expression for a given pressure P .

For example, for a pressure $P = 4$ bar and temperature $T = 105$ °C we get the following values:

$$K_{methane} = 0.1388 \quad (36)$$

$$K_{ethane} = 0.0001583 \quad (37)$$

$$K_{propane} = 1.779 \times 10^{-6} \quad (38)$$

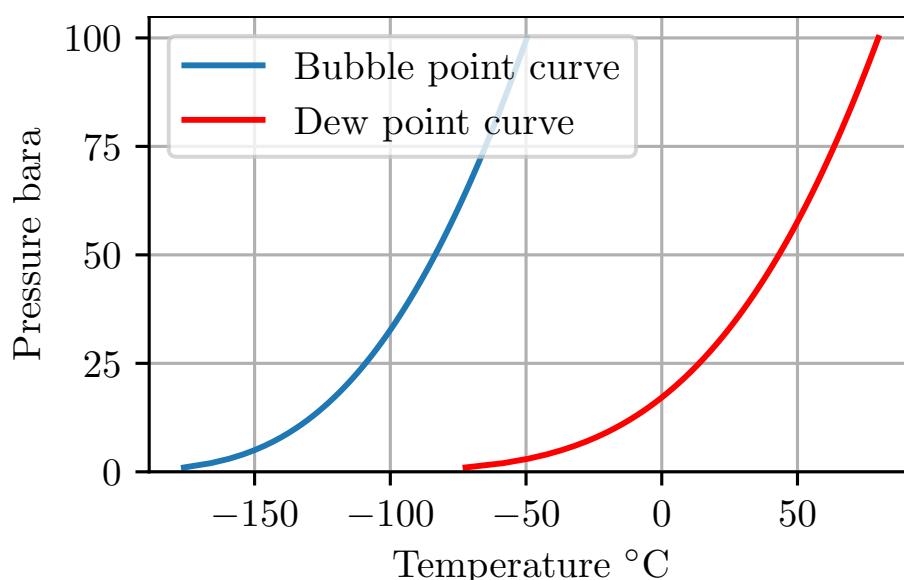
$$K_{i-butane} = 9.744 \times 10^{-8} \quad (39)$$

$$K_{n-butane} = 2.761 \times 10^{-8} \quad (40)$$

Finally:

$$\sum_i x_i K_i = 0.09024 \quad (41)$$

Solving for the value which satisfies Eq. 35 gives a temperature $T \approx 251$ K. Repeat this calculation for a range of pressure values in order to obtain the bubble point curve.



Note that the bubble and dew point curves will not converge at the critical point using this method. The Wilson expression is only applicable up to $P \approx 40$ bara.

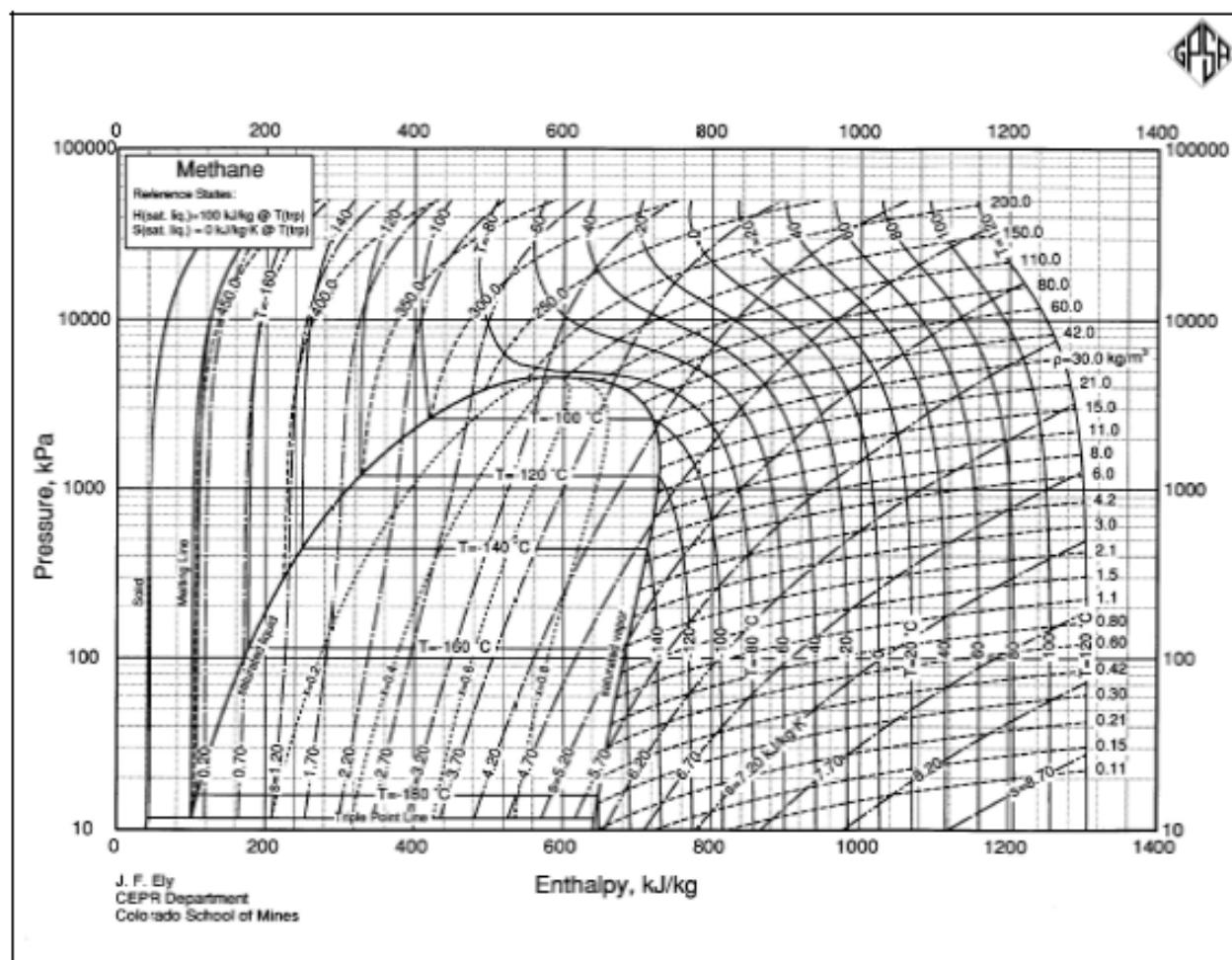
[Question end]

Q.4

Question 4

Methane is expanded across a choke valve from 100 bara and 20 °C to 1 bara. Using the attached Mollier chart calculate the downstream temperature. [-26 °C]

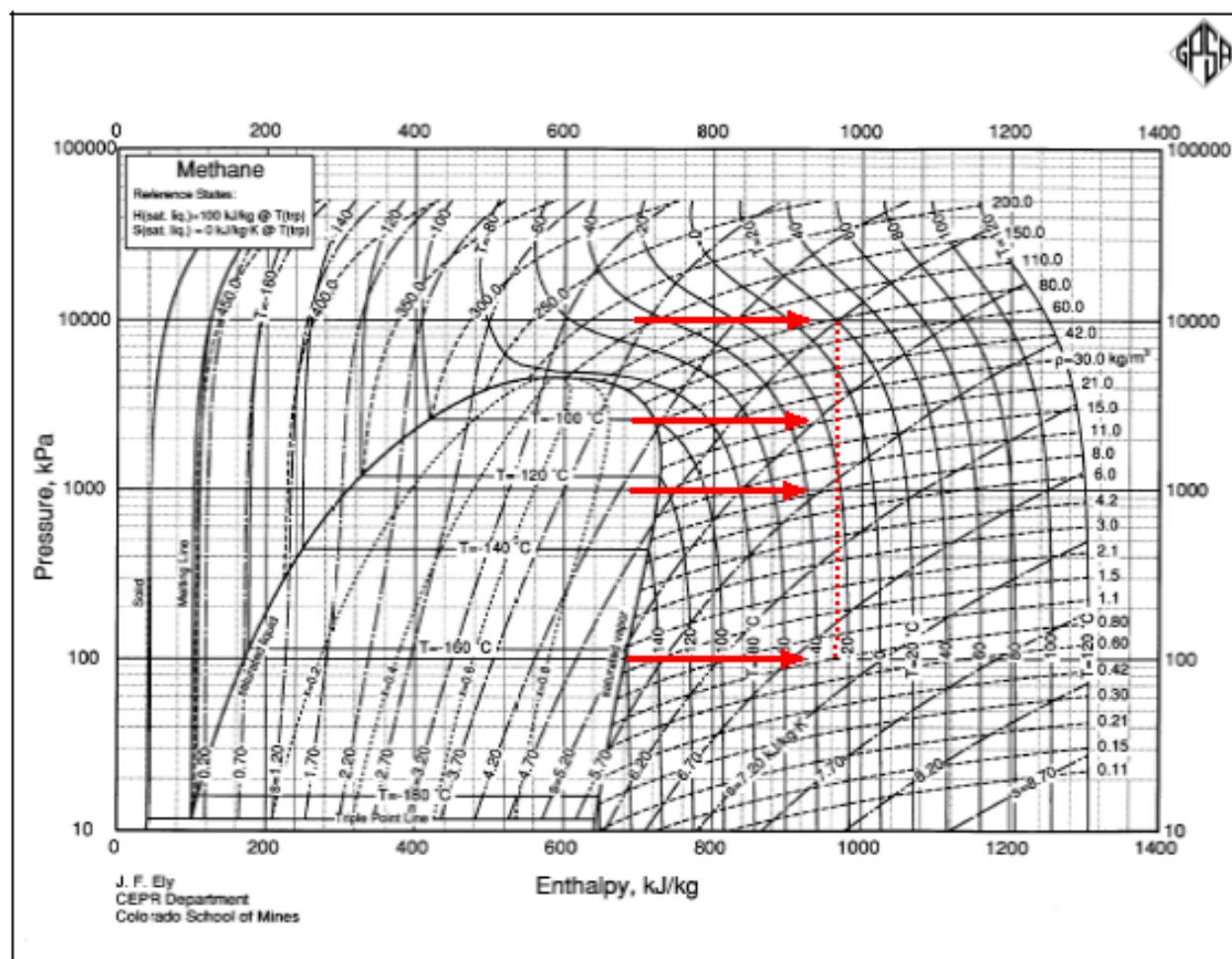
As the pipeline fills with methane the back pressure increases. Calculate the choke downstream temperature at back pressures of 10 and 25 bara. [-20 °C, -18 °C]



Solution:

For expansion across a valve the process is isenthalpic - enthalpy remains constant.

Find the enthalpy of the methane using the initial pressure $P_0 = 100 \text{ bara}$ and temperature $T_0 = 20^\circ\text{C}$ and find the new temperature when expanded to $P = 1 \text{ bara} / 10 \text{ bara} / 25 \text{ bara}$.



This leads to the following temperatures:

| Pressure bara | Temperature °C |
|---------------|----------------|
| 100 | 20 |
| 25 | -18 |
| 10 | -20 |
| 1 | -26 |

[Question end]

Q.5

Question 5

Calculate the gas density for the following mixture at 139.4 bara and 331 K.

| Component | Mol fraction |
|------------------|--------------|
| N ₂ | 0.0046 |
| CO ₂ | 0.0030 |
| H ₂ S | 0.1438 |
| Methane | 0.8414 |
| Ethane | 0.0059 |
| Propane | 0.0008 |
| i-Butane | 0.0003 |
| n-Butane | 0.0002 |

State your data sources. [120 kg m⁻³]

Solution:

See example in lecture notes on physical properties. Obtain P_C , T_C , and M from literature.

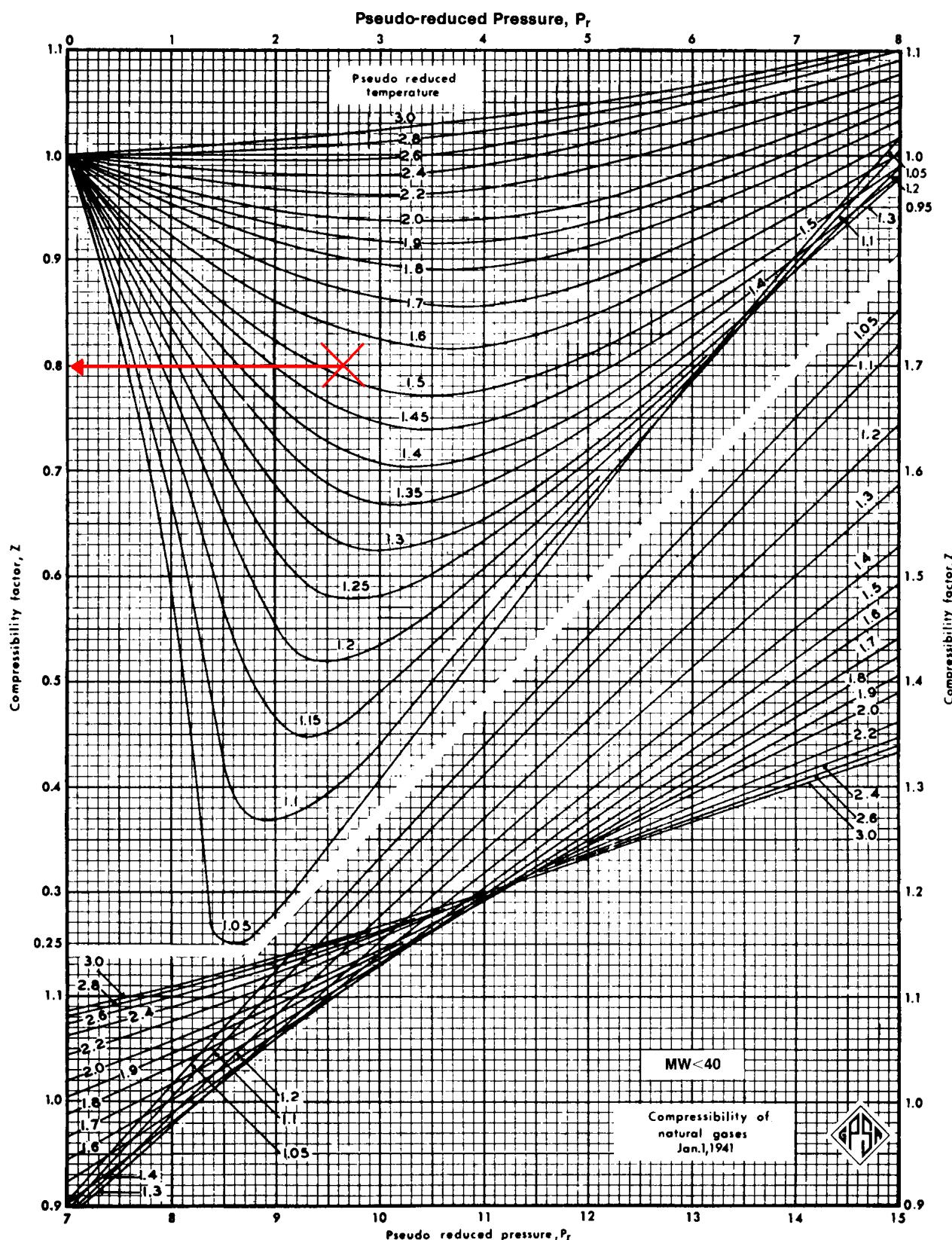
| Component | y_i | P_C | $P_C \cdot y_i$ | T_C | $T_C \cdot y_i$ | M | $M \cdot y_i$ |
|------------------|--------|-------|-----------------|-------|-----------------|-------|---------------|
| N ₂ | 0.0046 | 3.40 | 0.01564 | 126 | 0.6 | 28.02 | 0.13 |
| CO ₂ | 0.003 | 7.38 | 0.02214 | 304 | 0.9 | 44.01 | 0.13 |
| H ₂ S | 0.1438 | 8.96 | 1.28845 | 373 | 53.6 | 34.08 | 4.90 |
| Methane | 0.8414 | 4.60 | 3.87044 | 191 | 160.7 | 16.04 | 13.50 |
| Ethane | 0.0059 | 4.88 | 0.02879 | 305 | 1.8 | 30.07 | 0.18 |
| Propane | 0.0008 | 4.25 | 0.00340 | 370 | 0.3 | 44.10 | 0.04 |
| i-Butane | 0.0003 | 3.65 | 0.00110 | 408 | 0.1 | 58.12 | 0.02 |
| n-Butane | 0.0002 | 3.80 | 0.00076 | 425 | 0.1 | 58.12 | 0.01 |
| sum | 1 | | 5.23 Mpa | | 218 K | | 18.90 |

Calculate reduced pressure and temperature for overall mixture:

$$P_r = \frac{P}{\sum_i P_{Ci} y_i} = \frac{139.4}{52.3} = 2.67 \quad (42)$$

$$T_r = \frac{T}{\sum_i T_{Ci} y_i} = \frac{331}{218} = 1.52 \quad (43)$$

Obtain the compressibility factor $Z = 0.8$ from compressibility chart below.



Calculate the density:

$$\rho_m = \frac{PM_m}{Z_m RT} = \frac{13.94 \cdot 18.90}{0.8 \cdot 0.00831 \cdot 331} = 120 \text{ kg m}^{-3} \quad (44)$$

[Question end]

Q.6

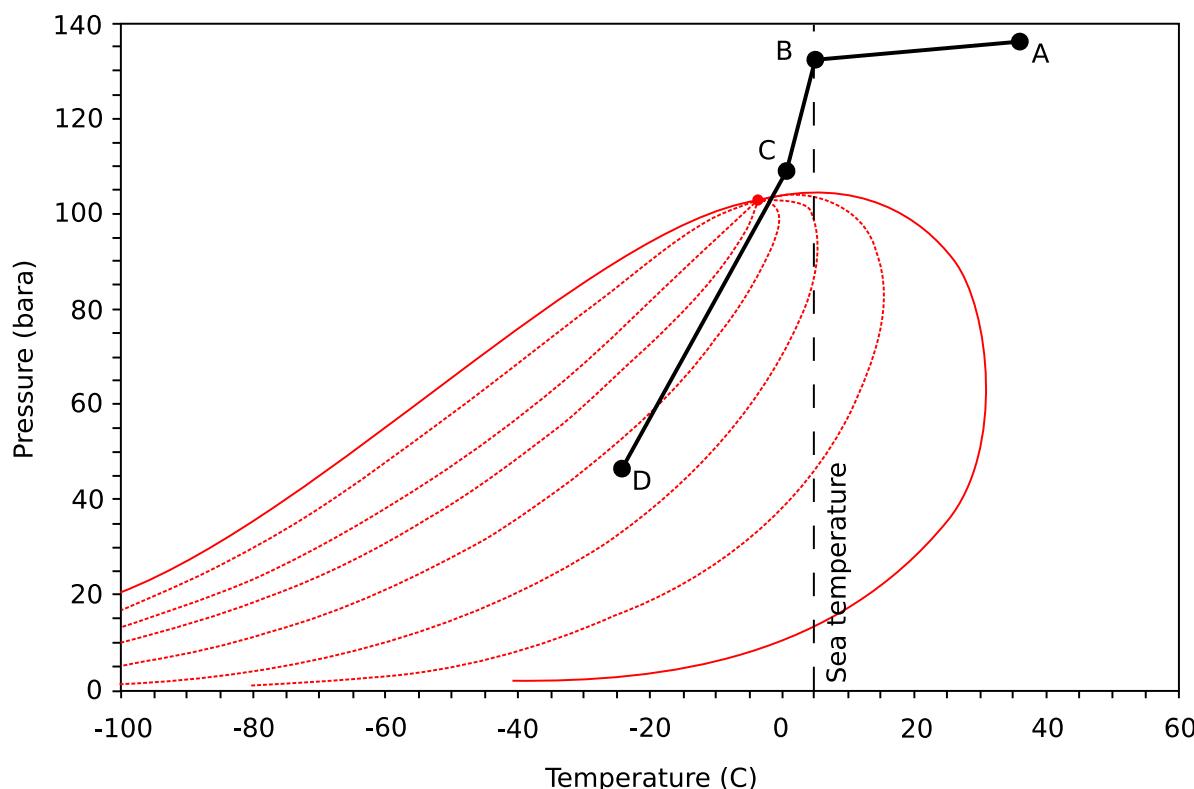
Question 6
Past exam question


Figure 1: The phase diagram for well-fluids produced offshore (point A) and the pressure-temperature path of the subsea pipeline ($A \rightarrow C$) and initial treatment at ($C \rightarrow D$) the processing plant.

Consider a subsea well. The wellhead temperature and pressure are indicated by point A on the figure above. The gas processing terminal, which is some distance away, receives the fluids at the conditions indicated by point C.

- a) Describe what might be occurring to cause the condition change $A \rightarrow B$. Note that the relative pressure drop is small. [5 marks]

Solution:

Pressure drop is proportional to pipe length, so we are seeing a fast cooling of the well fluids to the sea temperature upon initial entry into the pipeline. 5

- [5/5] b) Describe what might be occurring between points $B \rightarrow C$. In particular describe why the fluid is colder than the surrounding sea water. [5 marks]

Solution:

The (frictional/accelerational/hydrostatic) pressure drop along the pipeline is causing a Joule-Thomson effect which auto-refrigerates the fluid. This allows it to cool below the surrounding sea temperature. 5

- [5/5] c) Comment on the location of point C. Why is the subsea pipeline exit operated and designed at those conditions? Explain using key properties of the phase diagram. [5 marks]

Solution:

Point C is above the cricondenbar, which is the maximum pressure for the multi-phase

region of the phase envelope. The pipeline is thus operated in the dense-phase region of the phase diagram which removes any concern over multi-phase issues such as slugging or other intermittent flow as multiple phases cannot exist. It should be noted that this does not avoid issues such as entrained solids or other injected chemicals which are not accounted for in the phase envelope. 5

- [5/5] d) Upon arrival to the processing plant, the process C→D is carried out. Explain why this is performed and how the process might be carried out, particularly if energy use is to be minimized. [5 marks]

Solution:

The pressure is dropped upon entry to the plant to cause auto-refrigeration into the multi-phase region of the phase diagram. This allows a separation to “lean” the gas phase by knocking out the heavier, less volatile, hydrocarbons. This can be carried out using a Joule-Thomson valve, but for better energy efficiency a turbo-expander system would be used as the gas is typically recompressed for further transport. 5

[Question total: 20 marks]

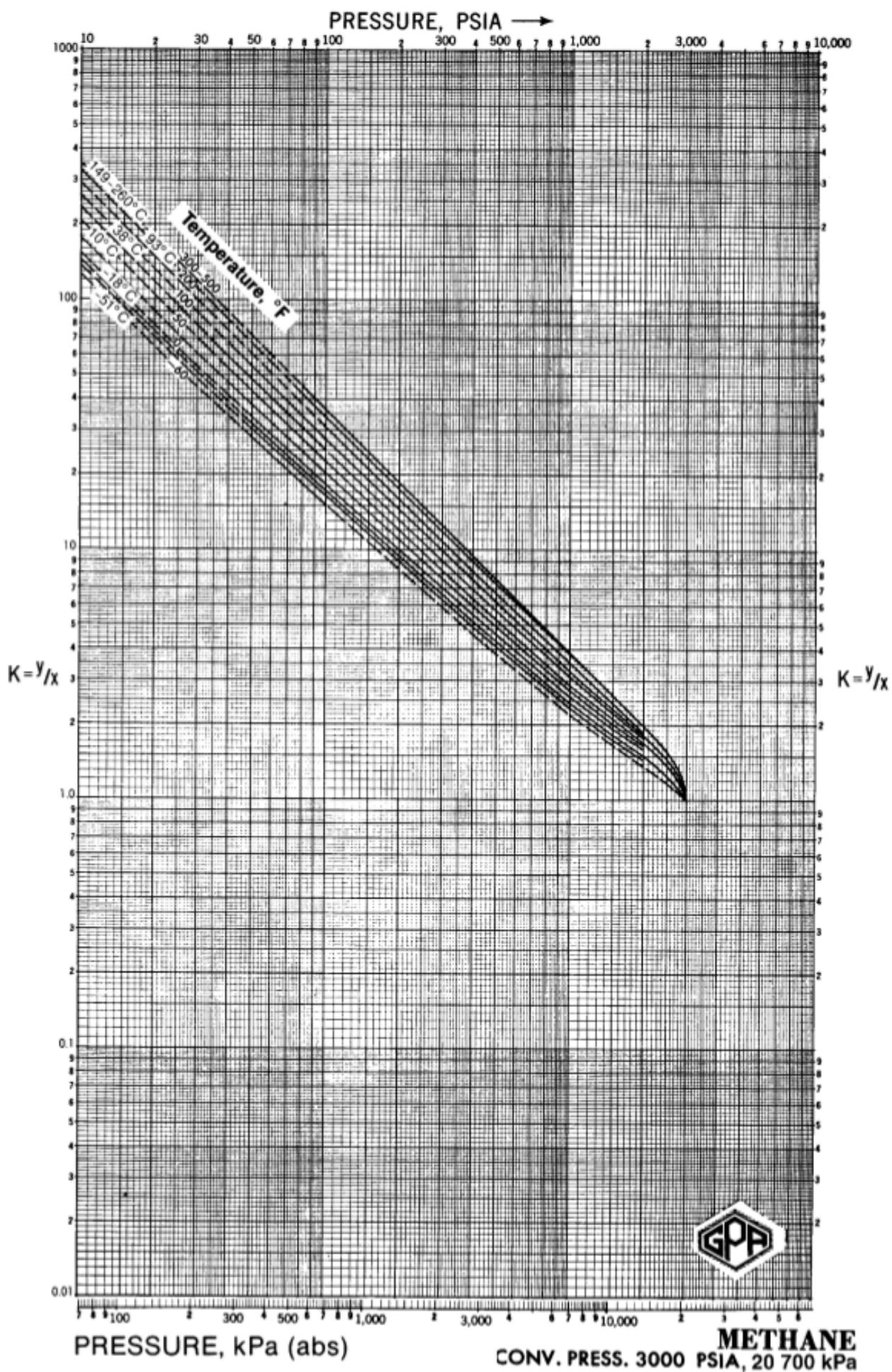
[5/5]

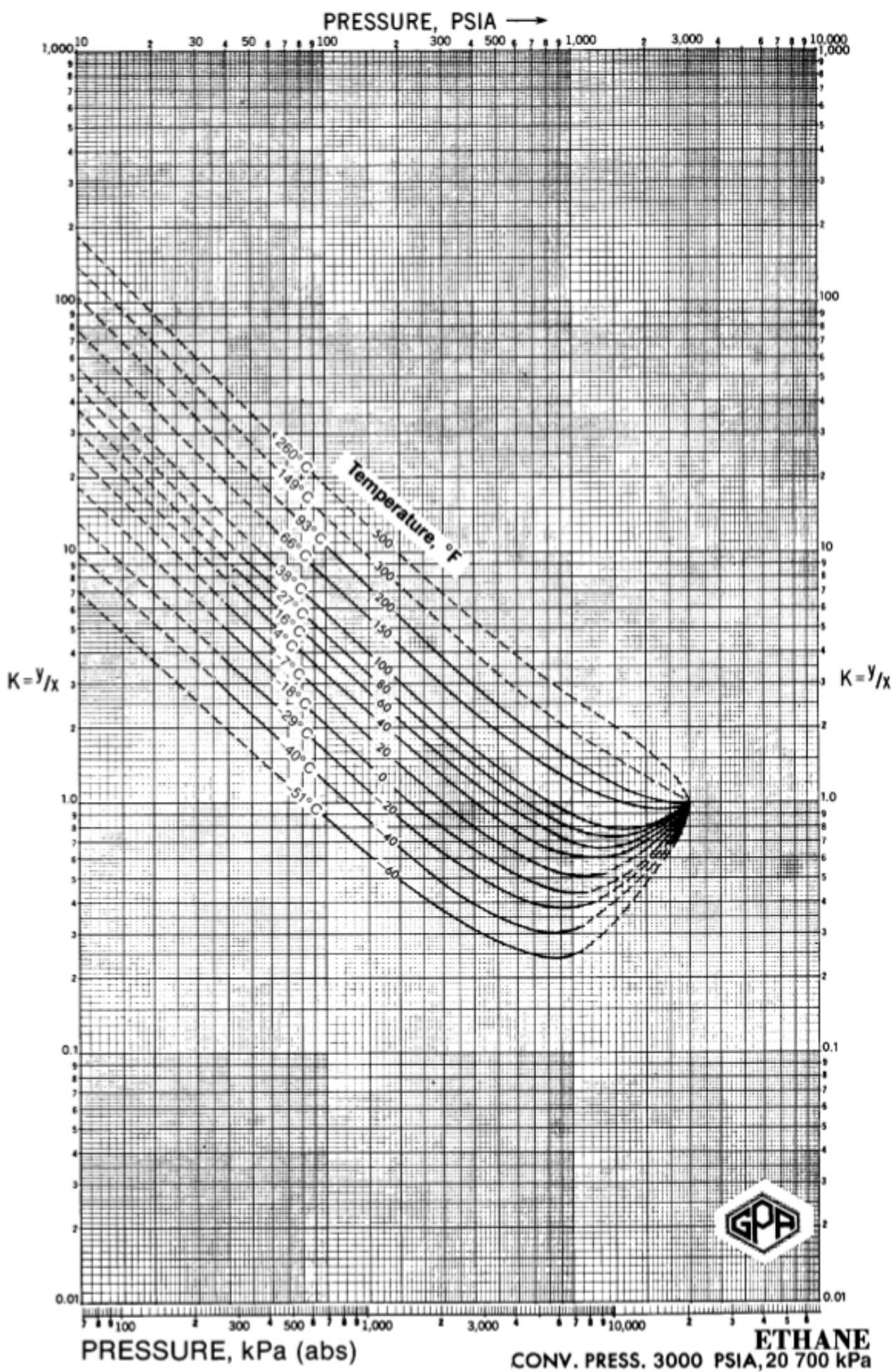
Q.7 Question 7

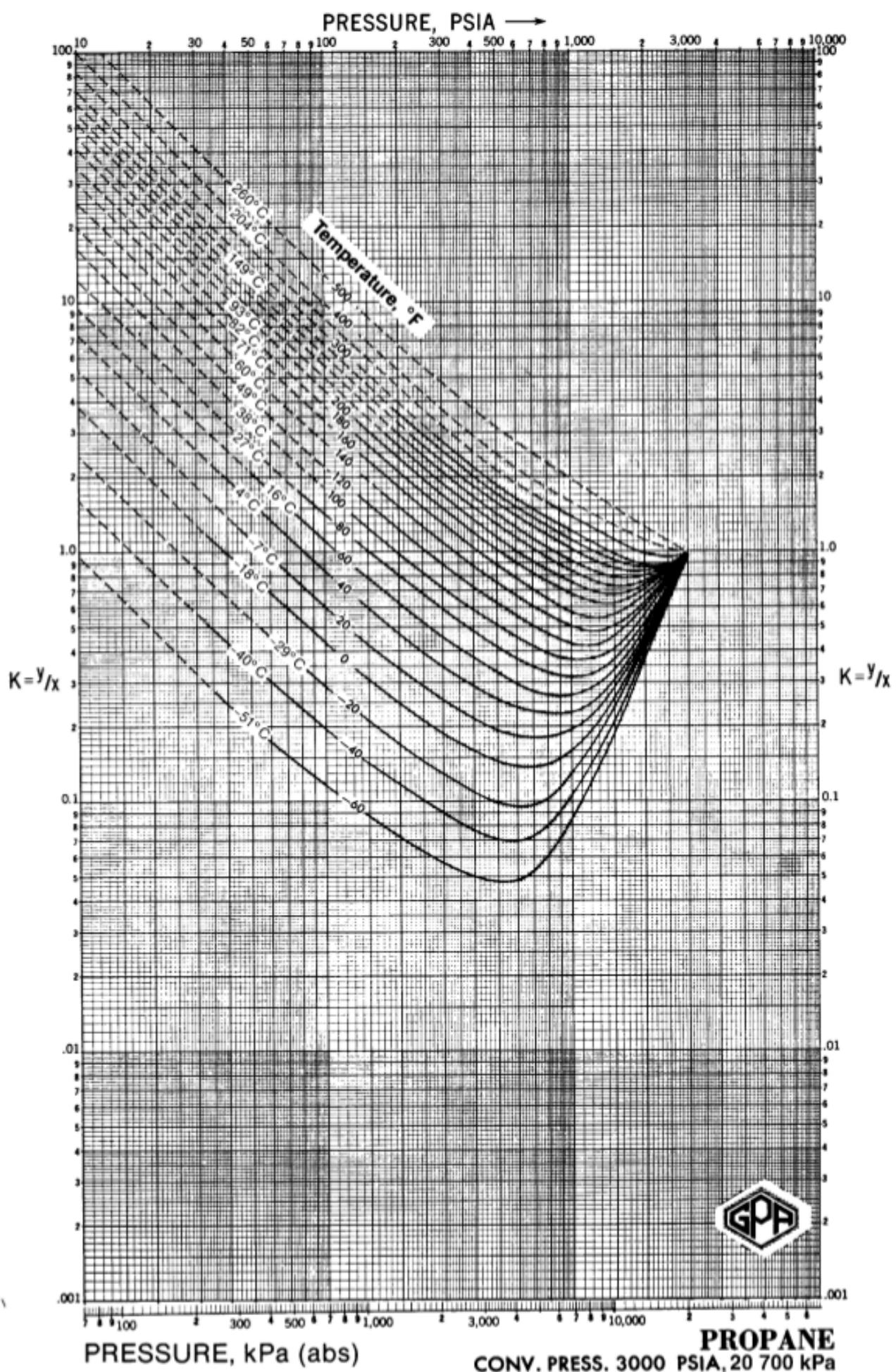
The following mixture flows in a subsea pipeline. The inlet temperature and pressure of the pipeline are 150 °F and 1000 psia the outlet conditions are 100 °F and 200 psia. Calculate the vapour liquid ratio at the pipeline inlet and outlet using the attached K value charts.

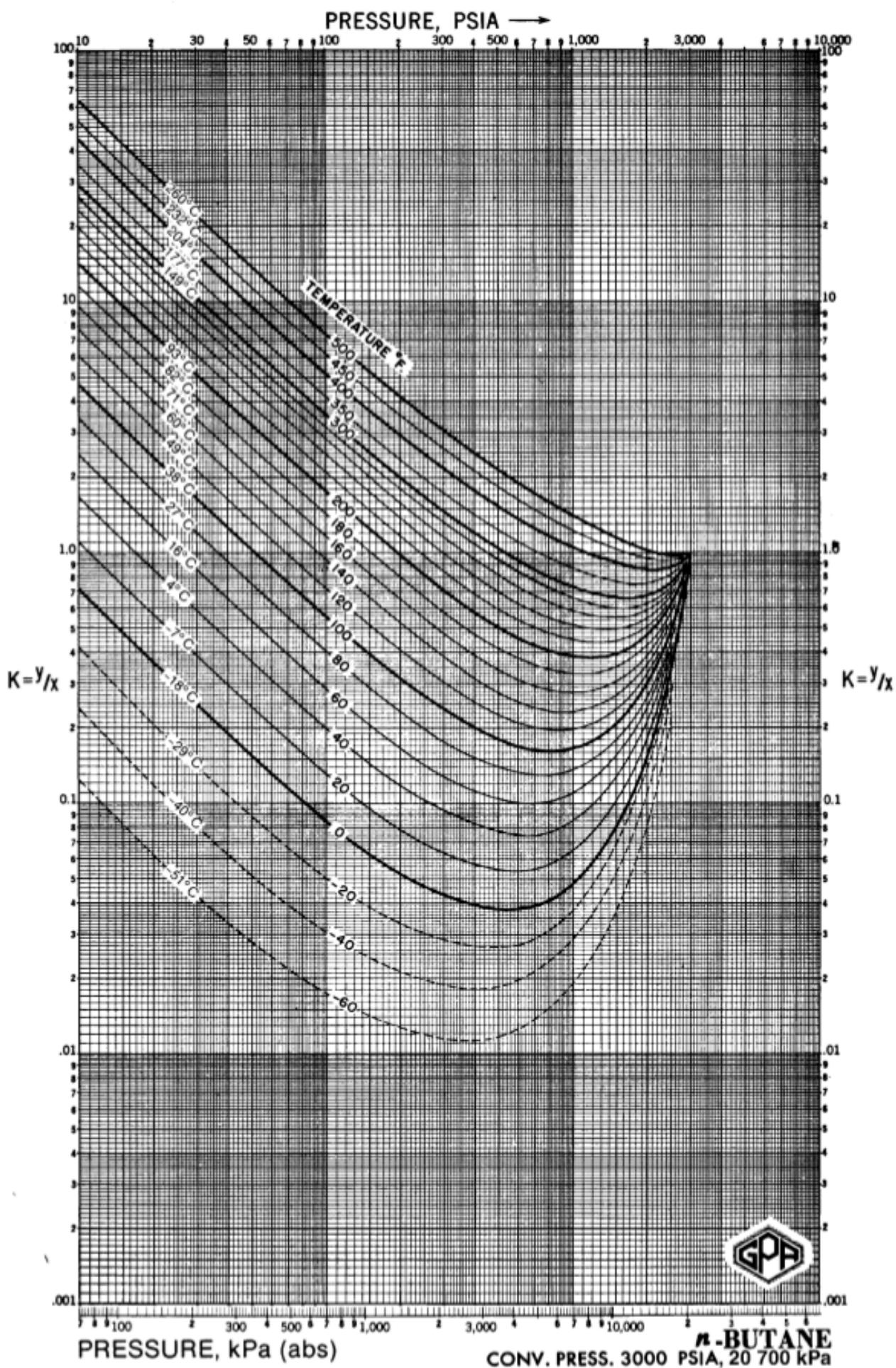
[Inlet V=0.14, L=0.86, Outlet V=0.67, L=0.33]

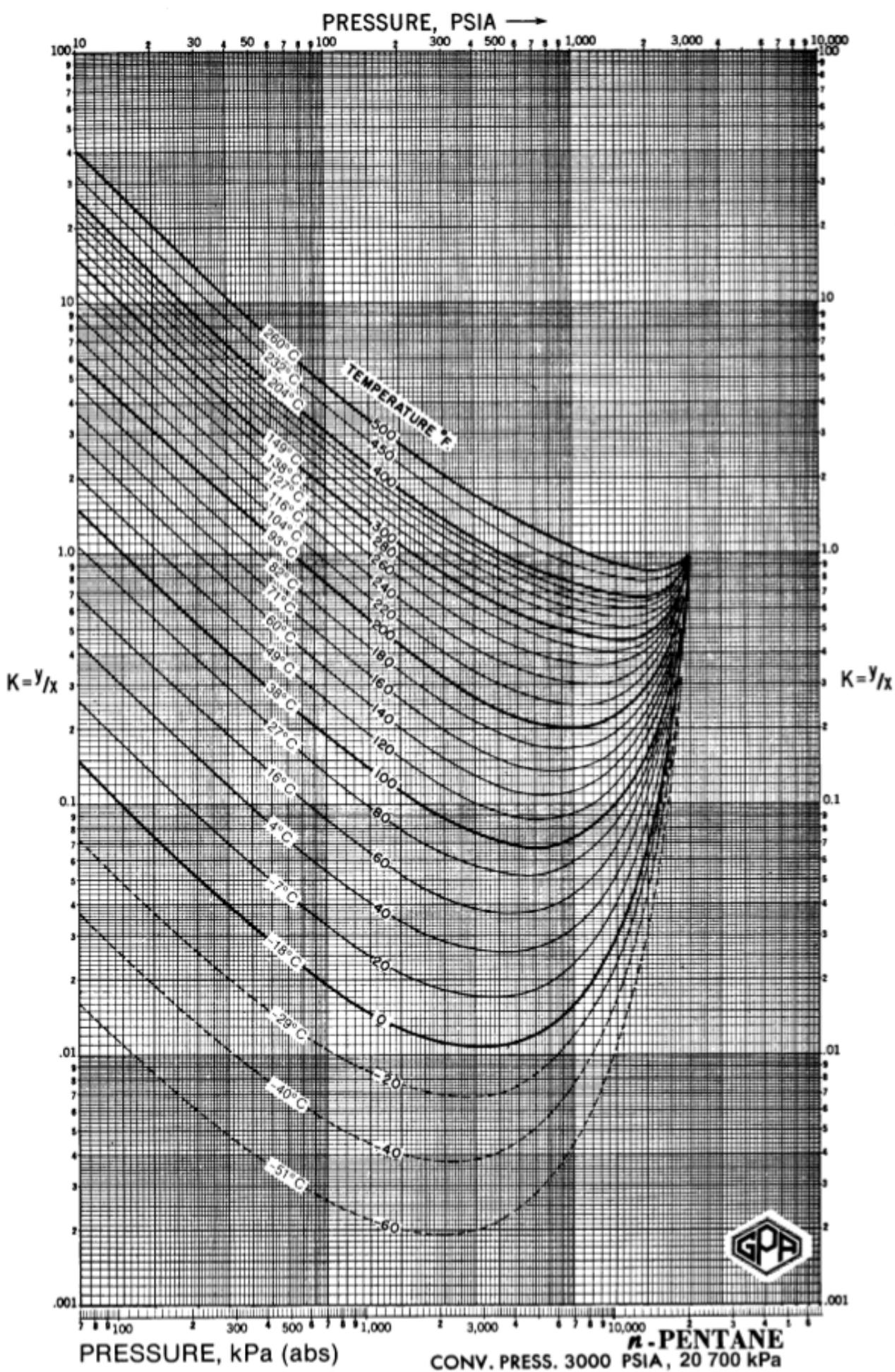
| Component | Mol fraction |
|-----------|--------------|
| Methane | 0.2 |
| Ethane | 0.3 |
| Propane | 0.15 |
| n-Butane | 0.1 |
| n-Pentane | 0.1 |
| n-Octane | 0.15 |

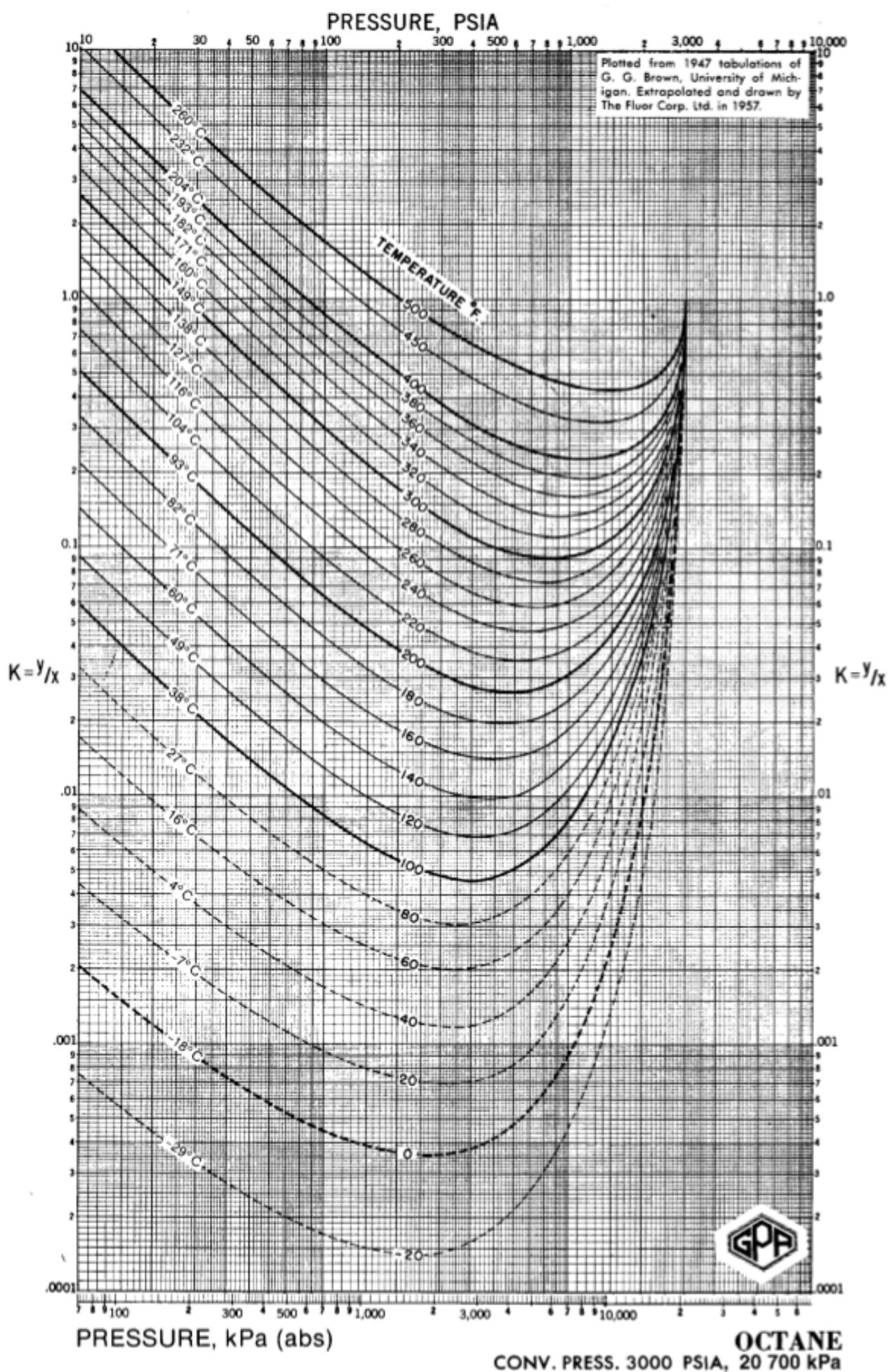












Solution:

Use the flash calculation procedure in lecture notes. Obtain K values from the charts for the pipeline inlet and outlet for each component. Begin with an initial guess for the volume fraction V and use this to calculate x_i values. Solve for the value of V that satisfies the expression

$$\sum_i x_i = 1.0 \quad (45)$$

Inlet: $V_{inlet} = 0.14$ and $L_{inlet} = (1 - V_{inlet}) = 0.86$.

| Component | Mol fraction | K_i | x_i | y_i |
|-----------|--------------|-------|----------|----------|
| Methane | 0.2 | 3.5 | 0.148148 | 0.518519 |
| Ethane | 0.3 | 1.2 | 0.291829 | 0.350195 |
| Propane | 0.15 | 0.52 | 0.160806 | 0.083619 |
| n-Butane | 0.1 | 0.23 | 0.112082 | 0.025779 |
| n-Pentane | 0.1 | 0.15 | 0.113507 | 0.017026 |
| n-Octane | 0.15 | 0.016 | 0.173965 | 0.002783 |
| | | Total | 1.000338 | 0.997921 |

Outlet: $V_{outlet} = 0.67$ and $L_{outlet} = (1 - V_{outlet}) = 0.33$.

| Component | Mol fraction | K_i | x_i | y_i |
|-----------|--------------|-------|----------|----------|
| Methane | 0.2 | 13 | 0.022124 | 0.287611 |
| Ethane | 0.3 | 4.2 | 0.09542 | 0.400763 |
| Propane | 0.15 | 1.5 | 0.11236 | 0.168539 |
| n-Butane | 0.1 | 0.6 | 0.136612 | 0.081967 |
| n-Pentane | 0.1 | 0.25 | 0.201005 | 0.050251 |
| n-Octane | 0.15 | 0.015 | 0.441112 | 0.006617 |
| | | Total | 1.008632 | 0.995748 |

[Question end]

Q.8**Question 8****1: Phase behaviour**

- a) Sketch a typical mixed hydrocarbon gas phase envelope. Indicate the cricondenbar. What is the significance of the Cricondenbar for the operation of gas pipeline systems? **[5 marks]**

Solution:

[2/5] Cricondenbar is the horizontal tangent at top of phase envelope.[✓]₂

[1/5] Irrespective of pipeline temperature, the gas can not enter the two-phase region if the pressure is above the cricondenbar[✓]₁.

[1/5] If liquid were to be produced, this would reduce the flow area, impacting production and could also cause troublesome slug flow[✓]₁. If a gas pipeline is operated above the cricondenbar, conditions promoting two-phase flow are prevented[✓]₁.

- [1/5]** b) An oil/gas separator is operating at 85°C and 2.3 bara producing an oil with a True Vapour Pressure (TVP) of 0.94 bara. Why is TVP an important specification? If water is introduced with the hydrocarbons, what will happen to the TVP if the separator pressure

and temperature remain unchanged?

[8 marks]

Solution:

It is important to maintain a TVP below atmospheric pressure to prevent vaporisation within the oil storage tanks. This could result in a loss of product and impact the safety of the storage tank by causing an over-pressure.[✓]

The TVP will reduce. The introduction of water produces an additional partial pressure of water onto the gas phase. Since the separator pressure is fixed the partial pressure associated with the gaseous hydrocarbon components will reduce. This effectively “boils off” more light ends resulting in a lowering of the liquid phase (oil) TVP.[✓]

- c) A subsea gas pipeline transports gas to a shore terminal. The seabed temperature is 4°C. The pipeline is being operated at a much higher gas throughput than normal and it is noted that the gas arrival temperature at the shore terminal is –2°C. What effect would explain a gas temperature lower than ambient temperature? [5 marks]

Solution:

At high gas rates, the pressure gradient will be correspondingly higher resulting in a more pronounced Joule Thomson expansion effect. The J-T expansion is reducing the gas temperature at a rate higher than the heat transfer rate from the sea is warming the gas.[✓]

- d) A natural gas has a calorific value of 40 MJ Nm⁻³. If the CO₂ concentration of the gas is increased from 0 to 5% calculate the new calorific value. [2 marks]

Solution:

CO₂ has no heat of combustion. Hence $4 \times 0.95 = 38$ MJ/Nm³.[✓]

[Question total: 20 marks]

Q.9

Question 9

1: Phase behaviour

- a) The following *T-p* plot is the phase envelope for a mixed hydrocarbon fluid. Describe the phase transitions moving from point A to B. [3 marks]

Solution:

The path is cooling at constant pressure. At point A the fluid is in the super-heated vapour state. Moving towards B the fluid cools and meets the dew-point line (red) and here the first droplets of liquid appear[✓]. As the fluid further cools more liquid is condensed. At the bubble point line (blue) the fluid is all liquid[✓]. Further liquid sub-cooling then occurs to point B₁[✓].

- b) The dew point of a mixed hydrocarbon gas can be calculated using the following expression:

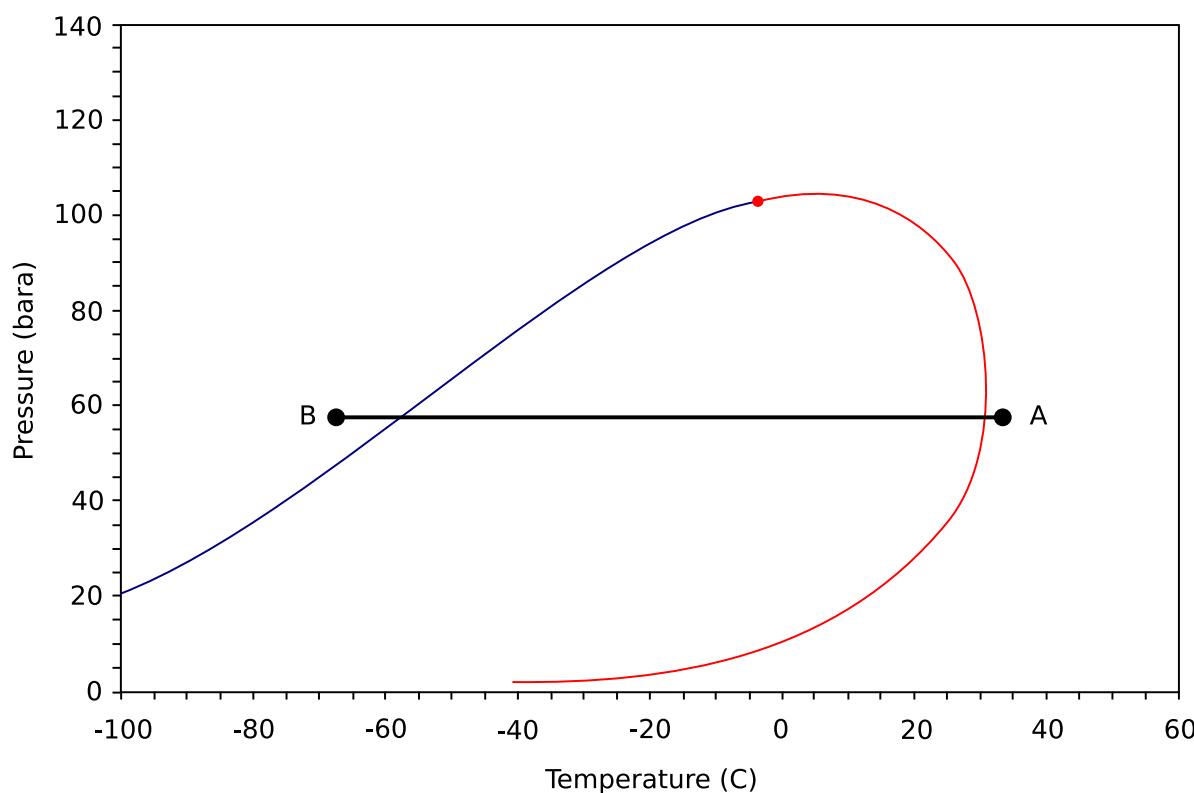
$$\sum_i \frac{y_i}{K_i} = 1.0$$

Describe the term K_i and how a typical K value chart such as that given in Fig. 2 may be used to calculate the dew point temperature at a fixed pressure of a multi-component mixture where the mole fractions are known. [6 marks]

Solution:

The K_i terms are equilibrium constants that are the ratio of component i mole fractions in the gas and liquid phase at equilibrium[✓].

A trial and error procedure is required. For a fixed pressure, a guess at the dew-point temperature is made. These are then used to look up the K_i values for each component from the chart[✓]. Using the equation above the sum is computed. If it does not equal 1, then the guessed temperature is revised until the sum converges to 1[✓].



- c) A final stage oil and gas separator is operating at 70 °C and 1.5 bara producing a crude oil product with a True Vapour Pressure (TVP) of 0.85 bara at 37.8 °C. Explain what will happen to the TVP of the crude if:
- the separator temperature reduces and the pressure is held constant.
 - the separator pressure reduces and the temperature is held constant. **[4 marks]**

Solution:

If the temperature reduces there will be less vapourisation of the light components hence the TVP of the crude oil will increase^{v2}.

If the pressure reduces there will be increased vapourisation of the light components hence the TVP of the crude oil will decrease^{v2}.

- d) Methane is expanded across a blowdown valve from 500 bara and 20 °C to 1.5 bara. Discuss the thermodynamics of the expansion process. Using the enclosed Mollier chart in Fig. 3 calculate the downstream temperature. **[3 marks]**

Solution:

The expansion across the valve is an isenthalpic process, hence follow the vertical line on a p - H chart from the initial conditions to 1.5 bara. Temperature will be approximately -90 °C^{v3}.

- e) A process simulation uses Peng-Robinson to determine the phase behaviour in the three-phase separator at a light crude offshore installation. Water is present in the received fluids, comment on the suitability/limitations of the results/model. **[4 marks]**

Solution:

The Peng-Robinson equation of state is a standard expression used widely in the upstream industry and should be suitable in general for this type of calculation; however, the Peng-

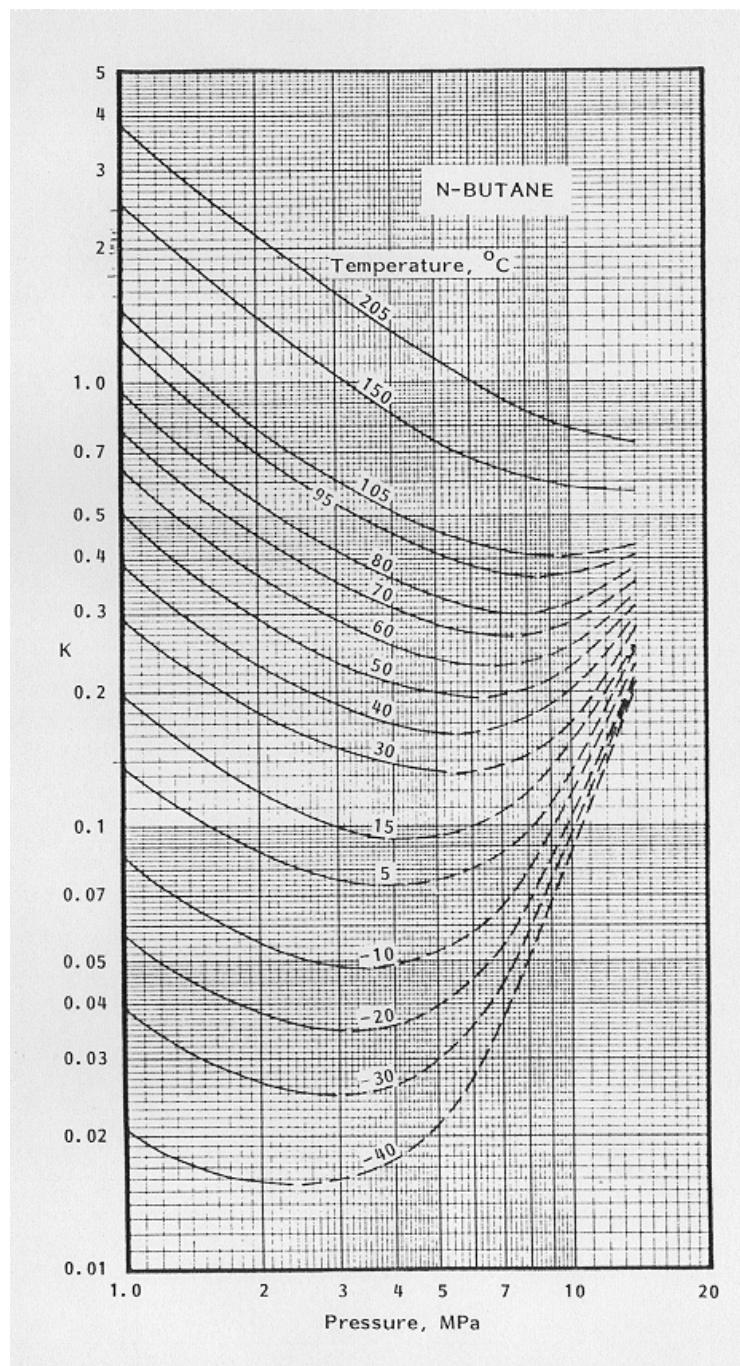


Figure 2: K-value chart for *n*-butane.

Robinson model will not correctly calculate the water partitioning but assume perfect oil-water separation[✓].

[4/4]

[Question total: 20 marks]

If you use this graph, you must attach it to your exam booklet using the provided tag.
Student ID:

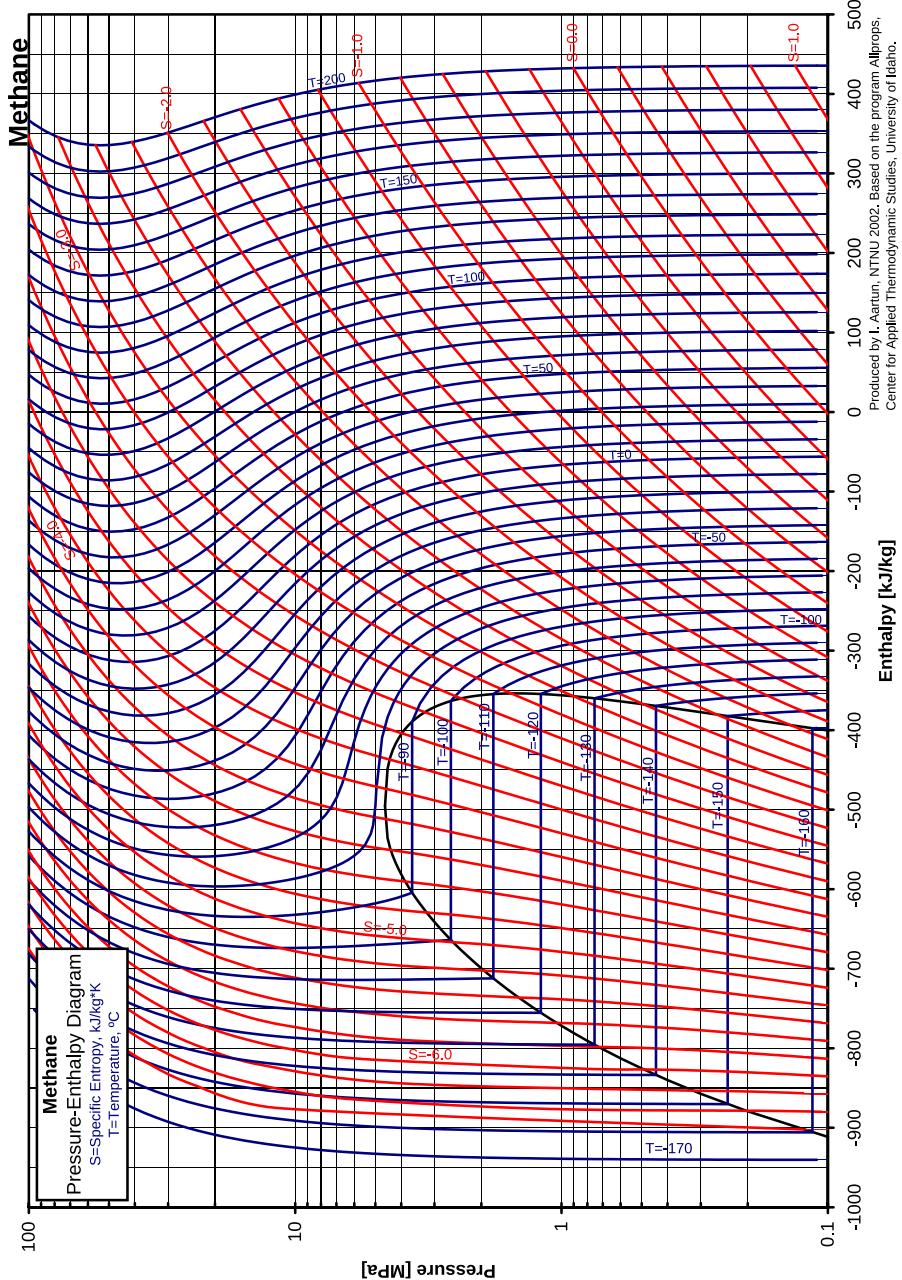


Figure 3: A Mollier chart for methane.

2 2: Fluid flow

Q.10

Question 10

An oil which has a viscosity of 50 cP ($0.05 \text{ kg m}^{-1} \text{ s}^{-1}$) and a density of 50 lb ft^{-3} (801.3 kg m^{-3}) flows in a 1 inch (25.4 mm) diameter pipe at the rate of $0.07 \text{ ft}^3 \text{ s}^{-1}$ ($0.00198 \text{ m}^3 \text{ s}^{-1}$). Determine the pressure drop which will occur over a 100 ft (30.48 m) section of the pipe. Undertake calculation for both imperial and metric units. [42.8 psi, 295.5 kPa]

Solution:

Firstly, calculate the Reynolds number to see which formula applies here. We calculate the fluid velocity inside the pipe

$$u = \frac{Q}{A} = \frac{4Q}{\pi d^2} = \frac{4 \cdot 0.00198}{\pi \cdot 0.0254^2} = 3.91 \text{ m s}^{-1} \quad (46)$$

which allows us to calculate the Reynolds number

$$Re = \frac{\rho u v}{\mu} = \frac{801.3 \cdot 3.91 \cdot 0.0254}{0.05} = 1591 \quad (47)$$

Since $Re < 2000$, this is laminar flow. We therefore calculate the friction factor using the following expression

$$f = \frac{64}{Re} \quad (48)$$

The pressure drop is calculated using Darcy's law

$$\Delta p = f \frac{\rho u^2 L}{2d} \quad (49)$$

$$= \frac{64}{1591} \frac{801.3 \cdot 3.91^2 \cdot 30.48}{2 \cdot 0.0254} \quad (50)$$

$$= 295.6 \text{ kPa} \quad (51)$$

[Question end]

Q.11 Question 11

A 160 km long subsea gas pipeline has an internal diameter of 333 mm. The inlet pressure is 90 bara and the outlet pressure is 20 bara. The pipeline flows isothermally at 4 °C. The gas consists of 75 % methane, 21 % ethane and 4 % propane. The gas compressibility is 0.85. Calculate the flowrate of gas in $\text{sm}^3 \text{ day}^{-1}$. The pipeline efficiency is 1. Compare the Weymouth and Panhandle A equations.

[3.13 million $\text{sm}^3 \text{ day}^{-1}$ Weymouth, 4.18 million $\text{sm}^3 \text{ day}^{-1}$ Panhandle A]

Solution:**Weymouth**

$$q_{SC} = K \left(\frac{T_{SC}}{P_{SC}} \right)^{1.000} \left(\frac{(P_1^2 - P_2^2) D^{5.333}}{\gamma L T_m Z_m} \right)^{0.5} E \text{ where } K = 1.185 \times 10^7 \quad (52)$$

Gas specific gravity $\gamma = \frac{\rho_{gas}}{\rho_{air}} = \frac{0.75 \cdot 16 + 0.21 \cdot 30 + 0.04 \cdot 44}{29} \approx 0.69172$.

$$q_{SC} = 1.185 \times 10^7 \left(\frac{288}{1 \times 10^5} \right)^{1.000} \left(\frac{\left((90 \times 10^5)^2 - (20 \times 10^5)^2 \right) 0.333^{5.333}}{0.69172 \cdot 160000 \cdot 277 \cdot 0.85} \right)^{0.5} 1.0 \quad (53)$$

$$= 3125929 \text{ m}^3 \text{ day}^{-1} \quad (54)$$

Panhandle A

$$q_{SC} = K \left(\frac{T_{SC}}{P_{SC}} \right)^{1.0788} \left(\frac{(P_1^2 - P_2^2) D^{4.854}}{\gamma^{0.8541} L T_m z_m} \right)^{0.5394} E \text{ where } K = 1.364 \times 10^7 \quad (55)$$

$$= 4180812 \text{ m}^3 \text{ day}^{-1} \quad (56)$$

[Question end]

Q.12 Question 12

If the gas entered the pipeline at 60 °C and cooled to 4 °C, outline how you would calculate the pressure drop in the pipeline if the gas flowrate was fixed.

Solution:

The flow is no longer isothermal, therefore a heat balance is required to predict the pipeline temperature profile. The temperature will affect the physical properties and the calculation should be conducted in multiple isothermal increments.

[Question end]

Q.13 Question 13

A square edged orifice of 50 mm diameter is installed in a 102.5 mm inside diameter pipe having a mercury manometer connected between the pipe taps. Find the flowrate of water when the mercury deflection is 110 mm. [395.3 litre min⁻¹]

Solution:

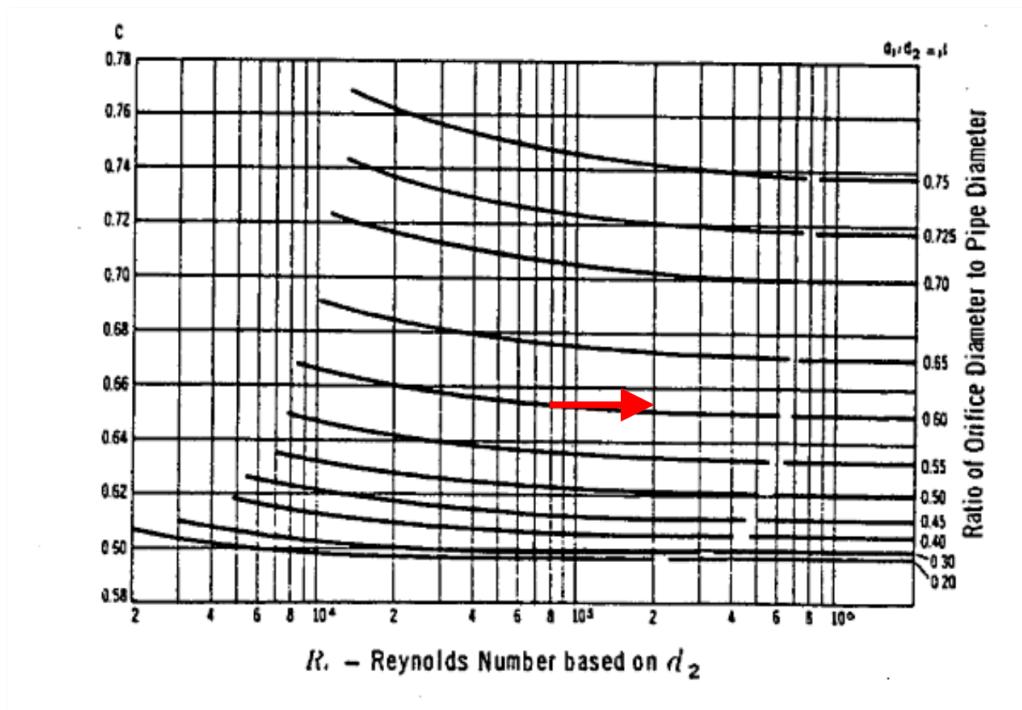
We will calculate the flowrate through the orifice using the following equation (from lecture notes)

$$q = CA_{orifice} \sqrt{\frac{2\Delta p}{\rho}} \quad (57)$$

Firstly the pressure gradient from the mercury reading is calculated (converting mm Hg to Pa: 0.75 mm Hg = 10.2 mm H₂O = 100 Pa)

$$\Delta p = \frac{110}{0.75} \times 100 = 14667 \text{ Pa} \quad (58)$$

Find a value of C using the figure below (see lecture notes) using $\frac{d_1}{d_2} = 0.4878 \approx 0.5$ and assuming turbulent flow (i.e. Re >> 2000)



$C = 0.62$, therefore

$$q = CA_{\text{orifice}} \sqrt{\frac{2\Delta p}{\rho}} = 0.62 \cdot \frac{\pi 0.05^2}{4} \cdot \sqrt{\frac{214667}{1000}} = 0.00659 \text{ m}^3 \text{ s}^{-1} \quad (59)$$

Now, all that remains is to check whether the initial assumption of turbulent flow was valid by calculating the Reynolds number (based on d_2 as the graph above indicates),

$$\text{Re}_{d_2} = \frac{\rho u d_2}{\mu} = \frac{\rho q d_2}{A \mu} = \frac{4 \rho q}{\pi d_2 \mu} = \frac{4 \cdot 1000 \cdot 0.00659}{\pi \cdot 0.1025 \cdot 0.0011} = 74418. \quad (60)$$

The assumption of turbulent flow was valid.

[Question end]

Q.14 Question 14

Using the following expression for frictional pressure drop

$$\frac{dp}{dl} = \frac{f \rho V^2}{2d} \quad (61)$$

and the ideal gas law develop an expression that takes the same form as the Panhandle A equation.

Solution:

$$q_{SC} = K \left(\frac{T_{SC}}{P_{SC}} \right)^{1.0788} \left(\frac{(P_1^2 - P_2^2) D^{4.854}}{\gamma^{0.8541} L T_m z_m} \right)^{0.5394} E \quad (62)$$

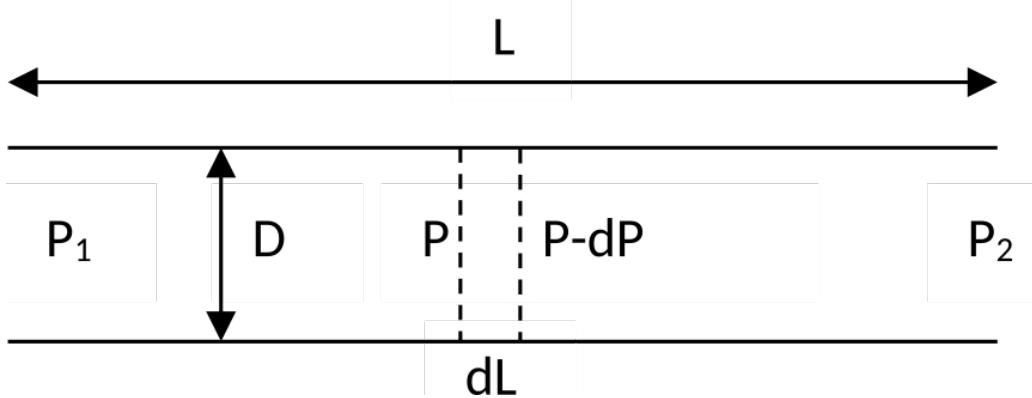
The pressure drop is given as

$$\frac{dp}{dl} = f \rho \frac{V^2}{2d} \quad (63)$$

and the ideal gas law is

$$V = \frac{nRTZ}{p} \quad (64)$$

sketch:



First we write the continuity equation

$$\rho_1 v_1 = \rho V \rightarrow V = \frac{\rho_1}{\rho} v_1 \quad (65)$$

We also recognise that density is proportional to pressure

$$\frac{p}{\rho} = \frac{p_1}{\rho_1} \rightarrow \rho = \frac{p}{p_1} \rho_1 \quad (66)$$

which allows us to write

$$V = \frac{p_1}{p} v_1 \quad (67)$$

Substituting the expressions for ρ and V into Eq. 66 gives

$$\frac{dp}{dl} = \frac{f \rho v^2}{2d} \quad (68)$$

$$= \frac{p}{p_1} \rho_1 \left(\frac{p_1}{p} \right)^2 v_1^2 \frac{f}{2d} \quad (69)$$

$$= \frac{p_1}{p} \rho_1 v_1^2 \frac{f}{2d} \quad (70)$$

We use the ideal gas law to generate an expression for p_1

$$p_1 = \frac{n_1 RTZ}{V_1} = \frac{m_1 RTZ}{MV_1} = \rho_1 RTZ \quad (71)$$

where m is the mass, M is the molar mass, and R changes to the specific gas constant R/M . Substituting this into Eq. 70 gives

$$p \frac{dp}{dl} = \rho_1 RTZ \rho_1 v_1^2 \frac{f}{2d} \quad (72)$$

$$= RTZ \rho_1^2 v_1^2 \frac{f}{2d} \quad (73)$$

We can rewrite the continuity equation as

$$\rho_1 v_1 = \rho_{SC} v_{SC} = \rho_{SC} \frac{4Q_{SC}}{\pi d^2} \quad (74)$$

which when substituted into Eq. 73 gives

$$p \frac{dp}{dl} = RTZ \rho_{SC}^2 \left(\frac{4Q_{SC}}{\pi d^2} \right)^2 \frac{f}{2d} \quad (75)$$

Integrating from $l = 0$ and $p = p_1$ to $l = L$ and $p = p_2$

$$\int_{p_1}^{p_2} pdp = \int_0^L RTZ \rho_{SC}^2 \left(\frac{4Q_{SC}}{\pi d^2} \right)^2 \frac{f}{2d} dl \quad (76)$$

$$\frac{p_2^2 - p_1^2}{2} = \frac{8f}{\pi^2 d^5} \rho_{SC}^2 Q_{SC}^2 RTZL \quad (77)$$

and rearranging for Q_{SC} making use of $\rho_{SC} = p_{SC}/RT_{SC}$ gives

$$Q_{SC}^2 = 0.617 (p_2^2 - p_1^2) \frac{1}{f \rho_{SC} RTZL} d^5 \quad (78)$$

$$= 0.785 \frac{1}{\rho_{SC}} \left(\frac{(p_2^2 - p_1^2) d^5}{f RTZL} \right)^{0.5} \quad (79)$$

$$= 0.785 \frac{T_{SC} R}{\rho_{SC}} \left(\frac{(p_2^2 - p_1^2) d^5}{f RTZL} \right)^{0.5} \quad (80)$$

[Question end]

Q.15

Question 15

The pressure at the top of a static gas filled riser is 20 bara, the temperature is 10 °C, the gas molecular weight is 22 g mol⁻¹, the compressibility at 20 bara is 0.88. The riser is 1000 m long, what is the pressure at the riser base? [22.2 bara]

Solution:

Use the hydrostatic gas equation

$$\frac{p_1}{\rho_1} \ln \left(\frac{p_2}{p_1} \right) + g(z_2 - z_1) = 0 \quad (81)$$

We first need to find ρ_1 .

$$\rho = \frac{Mn}{V} \quad (82)$$

where M is the molar mass. Use the ideal gas law to find n/V .

$$\frac{n}{V} = \frac{p}{RTZ} \quad (83)$$

Make sure that you use the correct dimensions for R , n (gm mol), T (K), p (kPa), and V (m³)

Values of the Gas Constant R in PV = nRT

Basis of units listed below is 22.4140 liters at 0°C and 1 atm for the volume of 1 g mole. All other values calculated from conversion factors listed in tables.

| n | Temperature | Pressure | Volume | R | n | Temperature | Energy | R |
|--------|-------------|--------------------|-----------------|---------------|--------|-------------|---------|--------------|
| gm mol | K | atm | liter | 0.082 057 477 | gm mol | K | calorie | 1.985 9 |
| gm mol | K | atm | cm ³ | 82.057 | gm mol | K | joule | 8.314 5 |
| gm mol | K | mm Hg | liter | 62.364 | | | | |
| gm mol | K | bar | liter | 0.083 145 | lb mol | °R | Btu | 1.985 9 |
| gm mol | K | kg/cm ² | liter | 0.084 784 | lb mol | °R | hp-hr | 0.000 780 48 |
| gm mol | K | kPa | m ³ | 0.0083145 | lb mol | °R | Kw-hr | 0.000 582 00 |
| lb mol | °R | atm | ft ³ | 0.730 24 | lb mol | °R | ft-lb | 1 545.3 |
| lb mol | °R | in.Hg | ft ³ | 21.850 | | | | |
| lb mol | °R | mm Hg | ft ³ | 554.98 | k mol | K | joule | 8 314.5 |
| lb mol | °R | lb/in ² | ft ³ | 10.732 | | | | |
| lb mol | °R | lb/ft ² | ft ³ | 545.3 | | | | |
| lb mol | K | atm | ft ³ | 1.3144 | | | | |
| lb mol | K | mm Hg | ft ³ | 998.97 | | | | |
| k mol | K | kPa | m ³ | 8.3145 | | | | |
| k mol | K | bar | m ³ | 0.083 145 | | | | |

$$\frac{n}{V} = \frac{p_1}{RT_1 Z} = \frac{2000}{0.0083415 \cdot 283 \cdot 0.88} \quad (84)$$

$$= 965.88 \text{ g mol m}^{-3} = 0.96588 \text{ kg mol m}^{-3} \quad (85)$$

Thus

$$\rho_1 = \frac{Mn}{V} = 22 \cdot 0.96588 = 21.25 \text{ kg m}^{-3} \quad (86)$$

Now use the hydrostatic equation to find p_2

$$\frac{2000 \cdot 10^3}{21.25} \ln \left(\frac{p_2}{2000 \cdot 10^3} \right) - 9.81 \cdot 1000 = 0 \quad (87)$$

therefore

$$p_2 = 2000 \cdot 10^3 \exp \left(\frac{9.81 \cdot 1000 \cdot 21.25}{2000 \cdot 10^3} \right) \quad (88)$$

$$= 2220 \text{ kPa} = 22.20 \text{ bara} \quad (89)$$

[Question end]

Q.16 Question 16

What is the riser base pressure if the riser from Q. 15 is full of oil with an sg of 0.825? [100.9 bara]

Repeat the calculation for a 600 m oil and 400 m gas filled riser. [69.4 bara]

Solution:

oil

$\gamma_{oil} = \rho_{oil}/\rho_{water}$. Therefore

$$\rho_{oil} = 0.825 \cdot 1000 = 825 \text{ kg m}^{-3} \quad (90)$$

$$\Delta p = \rho gh = 825 \cdot 9.81 \cdot 1000 = 8093250 \text{ Pa} \quad (91)$$

$$p_2 = p_1 + \Delta p = 20 \times 10^5 + 8093250 = 100.93 \text{ bara} \quad (92)$$

oil and gas

The first 400 m of the column will contain only gas as $\rho_{gas} < \rho_{oil}$. Therefore repeat Q. 15 using a length of 400 m.

$$p_2 = 2000 \cdot 10^3 \exp\left(\frac{9.81 \cdot 400 \cdot 21.25}{2000 \cdot 10^3}\right) \quad (93)$$

$$= 20.85 \text{ bara} \quad (94)$$

p_2 in this case is the pressure at the interface between the gas and the oil.

$$\Delta p_{oil} = \rho g h = 825 \cdot 9.81 \cdot 600 = 4855950 \text{ bara} \quad (95)$$

$$p_{base} = p_2 + \Delta p = 20.85 + 48.56 = 69.41 \text{ bara} \quad (96)$$

[Question end]

Q.17

Question 17

Water is flowing through a convergent divergent nozzle. Given the following information, calculate the pressure at the nozzle throat. Nozzle inlet pressure and temperature - 10 bara and 15 °C, inlet piping and nozzle diameter - 125 mm and 25 mm. The water flowrate is 1 m³ min⁻¹. Neglect frictional pressure drop. [4.245 bara]

Solution:

Use Bernoulli's equation

$$p_1 + \frac{\rho_1 v_1^2}{2} + \rho_1 g z_1 = p_2 + \frac{\rho_2 v_2^2}{2} + \rho_2 g z_2 \quad (97)$$

(in this case $z_1 = z_2$)

First, calculate the velocities

$$v_1 = \frac{q}{A_1} = \frac{4q}{\pi d_1^2} = \frac{4 \cdot 0.0166}{\pi \cdot 0.125^2} = 1.358 \text{ m s}^{-1} \quad (98)$$

$$v_2 = \frac{q}{A_2} = \frac{4q}{\pi d_2^2} = \frac{4 \cdot 0.0166}{\pi \cdot 0.025^2} = 33.953 \text{ m s}^{-1} \quad (99)$$

Calculate the nozzle pressure

$$p_2 = p_1 - \rho \left(\frac{v_2^2 - v_1^2}{2} \right) \quad (100)$$

$$= 10 \times 10^5 - 1000 \left(\frac{33.953^2 - 1.358^2}{2} \right) \quad (101)$$

$$= 424519 \text{ Pa} = 4.245 \text{ bara} \quad (102)$$

[Question end]

Q.18

Question 18

Calculate the speed of sound in air at 15 °C. [340 m s⁻¹]

Calculate the speed of sound in water. [1466 m s⁻¹]

Solution:

air

Calculate the speed of sound for an ideal gas

$$c = \sqrt{\frac{kRT}{M}} \quad (103)$$

where the molecular weight M of air is taken as 29 g mol^{-1} , and k_{air} is taken as 1.4 (from literature).

$$c = \sqrt{\frac{kRT}{M}} = \sqrt{\frac{1.4 \cdot 8314 \cdot 288}{29}} = 339.99 \text{ m s}^{-1} \quad (104)$$

water

The equation for the speed of sound in a liquid is

$$c = \sqrt{\frac{k}{\rho}} \quad (105)$$

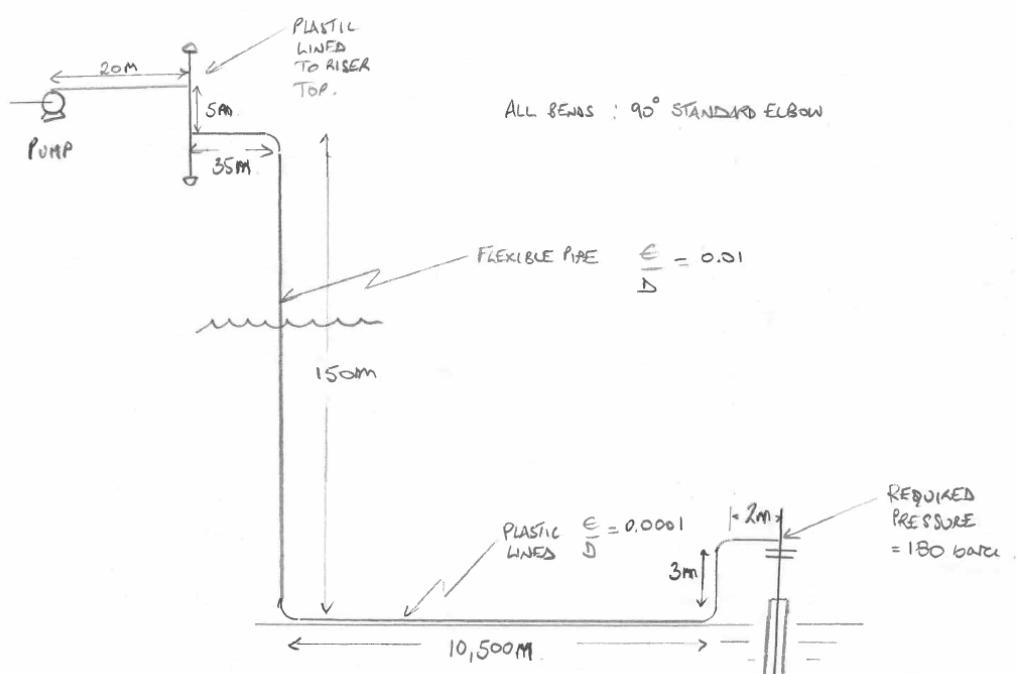
where the value for k_{water} is taken from the lecture notes.

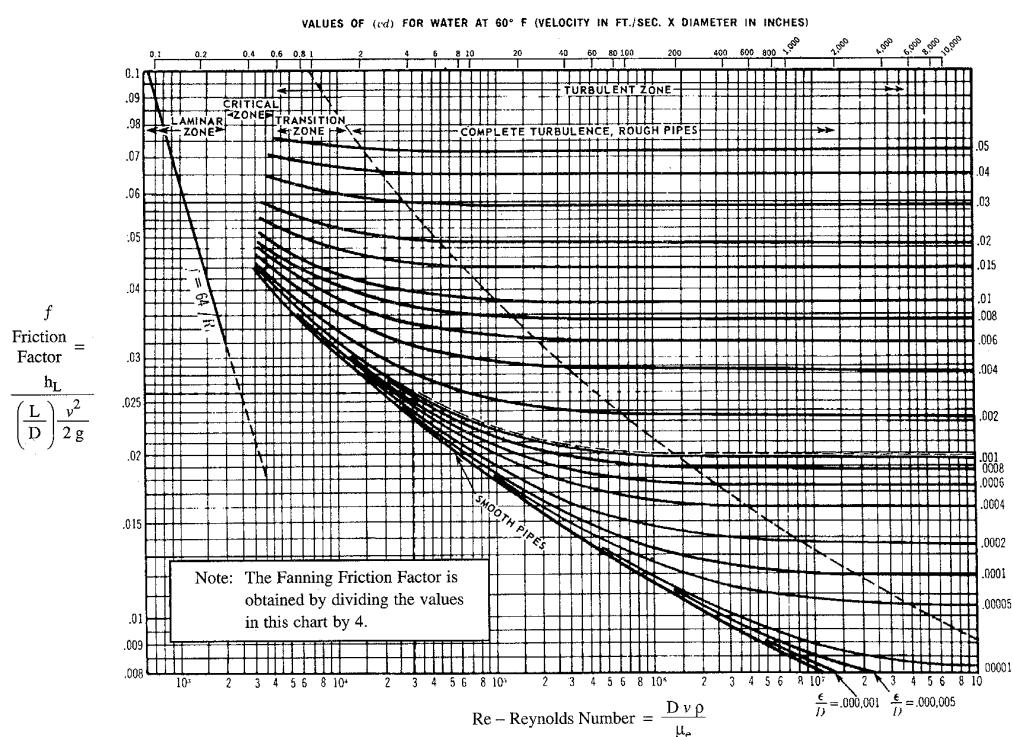
$$c = \sqrt{\frac{k}{\rho}} = \sqrt{\frac{2.15 \times 10^9}{1000}} = 1466.25 \text{ m s}^{-1} \quad (106)$$

[Question end]

Q.19 Question 19

A plastic lined, subsea water injection flowline is required to deliver $80,000 \text{ bbl day}^{-1}$ of treated sea water to a subsea manifold at a pressure of 180 bara and temperature of 10°C . The water pipeline route is as shown. The topsides piping is also plastic lined but the riser is flexible pipe. The pipe ID is 10.75 inch throughout. Calculate the required water injection pump discharge pressure. $6.29 \text{ bbl} = 1 \text{ m}^3$. Viscosity and density of sea water at 10°C is 1.31 cP and 1020 kg m^{-3} respectively. [185.0 bara]





| Description | | | | Equivalent Length in Pipe Diameters (L/D) |
|---|---|--|---|--|
| Globe Valves | Stem perpendicular to run | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 340 450 |
| | Y-pattern | Stem 60° from run of pipe Stem 45° from run of pipe | 100% Open 100% Open | 175 145 |
| Angle Valves | | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 145 200 |
| Gate Valves | Wedge Disc Double Disc, Plug Disc | | 100% Open 75% Open 50% Open 25% Open | 13 35 160 900 |
| | Pulp Stock | | 100% Open 75% Open 50% Open 25% Open | 17 50 260 1200 |
| Conduit PipeLine Gate, Ball and Plug Valves | | | | 100% Open 3 |
| Check Valves | Conventional Swing Clearway Swing Globe Lift or Stop; Stem Perpendicular to Run or Y-pattern Angle lift or Stop In-Line Ball | | | 135 50 same as globe same as globe 150 |
| Foot Valves with Strainer | | Poppet lift-type disc Leather-hinger disc | | 420 75 |
| Butterfly Valves (8-inch and larger) | | | | 40 |
| Cocks | Straight-through | Rectangular plug port area equal to 100% of pipe area | | 18 |
| | Three-way | Rectangular plug port area equal to 80% of pipe area | Straight through Through branch | 44 140 |
| Fittings | 90° Standard Elbow 45° Standard Elbow 90° Long Radius Elbow | | | 30 16 20 |
| | 90° Street Elbow 45° Street Elbow Square Corner Elbow | | | 50 26 57 |
| | Standard Tee | Flow through run Flow through branch | | 20 60 |
| | Close Pattern Return Bend | | | 50 |

Solution:

Pipeline diameter:

$$d = 10.72 \text{ in} = 0.2731 \text{ m} \quad (107)$$

Use equivalent length to calculate the pressure drop in elbows and branches. The pipeline contains:

- $2 \times$ Flow through branches ($60 \frac{L}{d}$). Therefore

$$L = 2 \cdot 60 \cdot 0.2731 = 32.8 \text{ m} \quad (108)$$

- $4 \times 90^\circ$ elbow ($30 \frac{L}{d}$). Therefore

$$L = 4 \cdot 30 \cdot 0.2731 = 32.8 \text{ m} \quad (109)$$

| Description | | | | Equivalent Length in Pipe Diameters (L/D) |
|--|---|--|---|--|
| Globe Valves | Stem perpendicular to run | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 340 450 |
| | Y-pattern | Stem 60° from run of pipe Stem 45° from run of pipe | 100% Open 100% Open | 175 145 |
| Angle Valves | | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 145 200 |
| Gate Valves | Wedge, Disc Double Disc, Plug Disc | | 100% Open 75% Open 50% Open 25% Open | 13 35 160 900 |
| | Pulp Stock | | 100% Open 75% Open 50% Open 25% Open | 17 50 260 1200 |
| Conduit Pipe Line Gate, Ball and Plug Valves | | | 100% Open | 3 |
| Check Valves | Conventional Swing Clearway Swing Globe Lift or Stop; Stem Perpendicular to Run or Y-pattern Angle lift or Stop In-Line Ball | | | 135 50 same as globe same as globe 150 |
| | | | | |
| Foot Valves with Strainer | | Poppet lift-type disc Leather-hinger disc | | 420 75 |
| Butterfly Valves (3-inch and larger) | | | | 40 |
| Cocks | Straight-through | Rectangular plug port area equal to 100% of pipe area | | 18 |
| | Three-way | Rectangular plug port area equal to 80% of pipe area | Straight through Through branch | 44 140 |
| Fittings | 90° Standard Elbow 45° Standard Elbow 90° Long Radius Elbow | | | 30 16 20 |
| | 90° Street Elbow 45° Street Elbow Square Comer Elbow | | | 50 26 57 |
| | Standard Tee | Flow through run Flow through branch | | 20 60 |
| | Close Pattern Return Bend | | | 50 |
| | | | | |
| | | | | |

The pipeline total equivalent length is

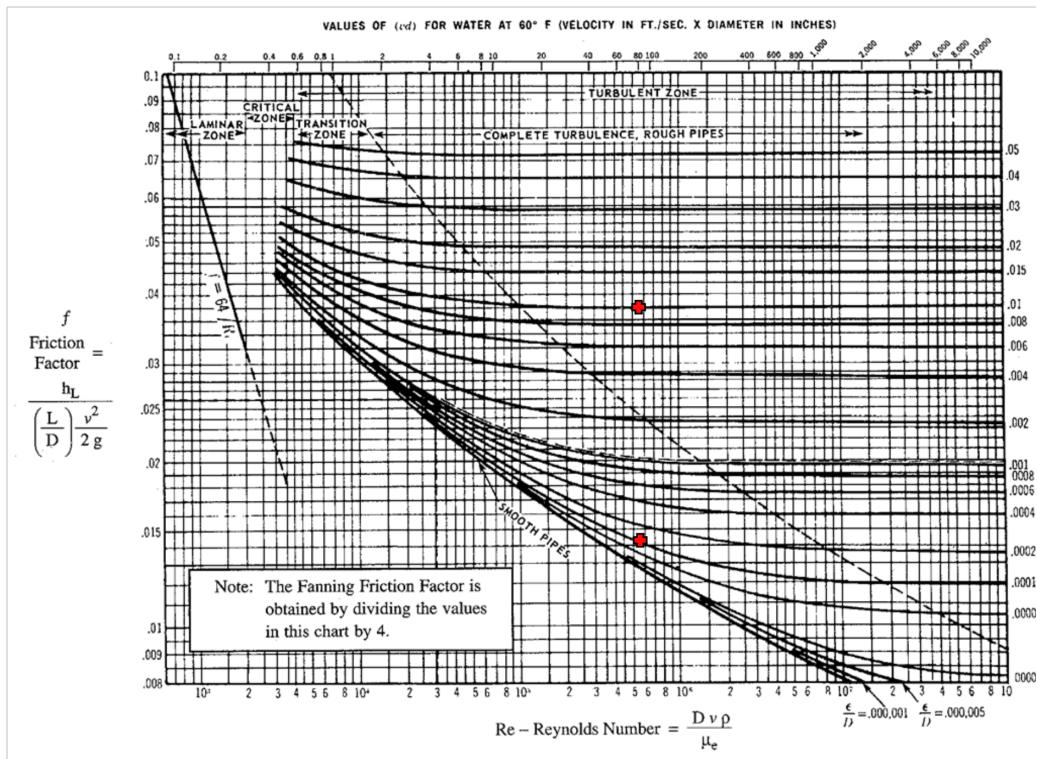
$$L = \underbrace{20 + 5 + 35 + 10500 + 3 + 2}_{\text{pipe length excluding riser}} + \underbrace{32.8 + 32.8}_{\text{equivalent lenght of fittings}} \quad (110)$$

$$= 10630 \text{ m} \quad (111)$$

Calculate f for both the plastic lined and flexible pipe. This requires us to calculate the Reynolds number. This calculation assumes a water viscosity at 10 °C of $\mu = 1.31 \text{ cP} = 0.00131 \text{ Pa s}$ and a water density at 10 °C of $\rho = 1020 \text{ kg m}^{-3}$

$$Re = \frac{\rho u d}{\mu} = \frac{1020 \cdot 2.54 \cdot 0.2731}{0.00131} \quad (112)$$

which allows us to use the Moody diagram to obtain f



- Flexible pipe $\rightarrow f = 0.037$
- Plastic lined pipe $\rightarrow f = 0.014$

We can now calculate the total frictional pressure drop in the plastic lined pipe

$$\Delta p = \frac{f \rho u^2 L}{2d} = \frac{0.014 \cdot 1020 \cdot 2.54^2 \cdot 10630}{2 \cdot 0.2731} = 18.6 \text{ bar} \quad (113)$$

and the flexible pipe

$$\Delta p = \frac{f \rho u^2 L}{2d} = \frac{0.037 \cdot 1020 \cdot 2.54^2 \cdot 150}{2 \cdot 0.2731} = 0.68 \text{ bar} \quad (114)$$

We now calculate the total elevation pressure drop. The total elevation change in this case is $\Delta z = 5 + 150 - 3 = 152 \text{ m}$.

$$\Delta p = \rho g z = 1020 \cdot 9.81 \cdot 152 = 15.2 \text{ bar} \quad (115)$$

We therefore require a pump discharge pressure of

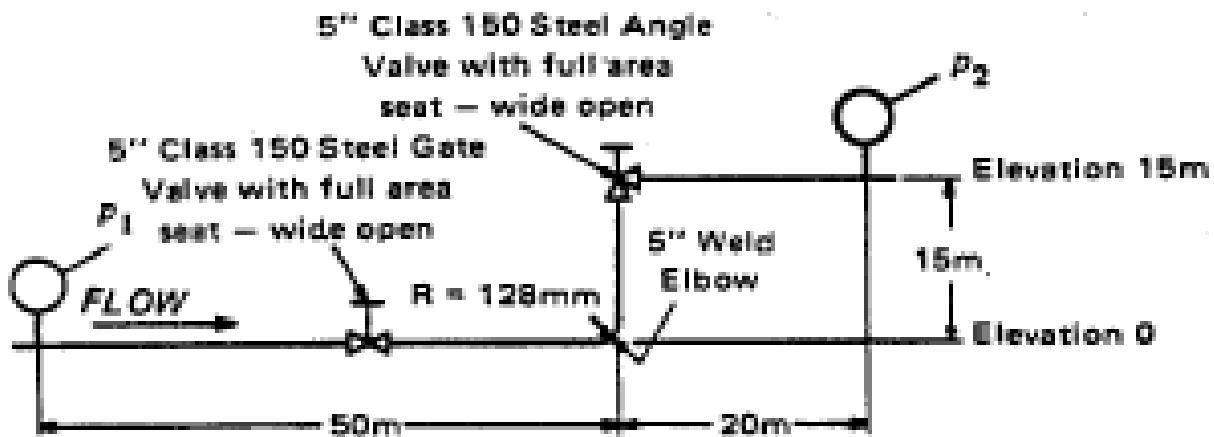
$$p_{pump} = 180 + 18.6 + 0.68 - 15.2 = 184.1 \text{ bar} \quad (116)$$

[Question end]

Q.20

Question 20

For the following piping system has an oil flow rate of 2300 l min^{-1} , pipe diameter of 128.2 mm, liquid density of 899 kg m^{-3} and viscosity of 450 cP. Calculate the pressure difference $p_1 - p_2$. The fully turbulent friction factor f_T is 0.016. For the fittings (valves/components) $\beta=1$. [3.65 bar]



APPENDIX A

CRANE PHYSICAL PROPERTIES OF FLUIDS AND FLOW CHARACTERISTICS OF VALVES, FITTINGS AND PIPE A - 27

"K" FACTOR TABLE – SHEET 2 of 4

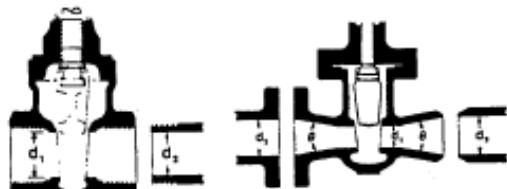
Representative Resistance Coefficients (K) for Valves and Fittings

(for formulas and friction data, see page A-26)

("K" is based on use of schedule pipe as listed on page 2-10)

GATE VALVES

Wedge Disc, Double Disc, or Plug Type



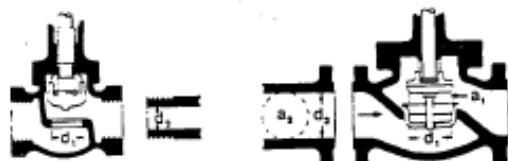
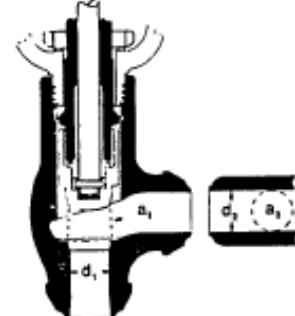
If: $\beta = 1, \theta = 0 \dots K_1 = 8 f_T$
 $\beta < 1 \text{ and } \theta < 45^\circ \dots K_1 = \text{Formula 5}$
 $\beta < 1 \text{ and } 45^\circ < \theta < 180^\circ \dots K_1 = \text{Formula 6}$

SWING CHECK VALVES



$K = 100 f_T$ $K = 50 f_T$
Minimum pipe velocity (mps) for full disc lift
 $= 45\sqrt{V}$ $= 75\sqrt{V} \text{ except}$
U/L listed = $120\sqrt{V}$

GLOBE AND ANGLE VALVES

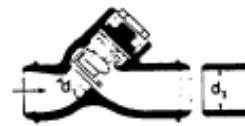
If: $\beta = 1 \dots K_1 = 340 f_T$ If: $\beta = 1 \dots K_1 = 55 f_T$ If: $\beta = 1 \dots K_1 = 150 f_T$ If: $\beta = 1 \dots K_1 = 55 f_T$ All globe and angle valves,
whether reduced seat or throttled,If: $\beta < 1 \dots K_1 = \text{Formula 7}$

LIFT CHECK VALVES



If: $\beta = 1 \dots K_1 = 600 f_T$
 $\beta < 1 \dots K_1 = \text{Formula 7}$

Minimum pipe velocity (mps) for full disc lift
 $= 50\beta^2 \sqrt{V}$



If: $\beta = 1 \dots K_1 = 55 f_T$
 $\beta < 1 \dots K_1 = \text{Formula 7}$

Minimum pipe velocity (mps) for full disc lift
 $= 170\beta^2 \sqrt{V}$

TILTING DISC CHECK VALVES



| Sizes | $\alpha = 5^\circ$ | $\alpha = 15^\circ$ |
|--|--------------------|---------------------|
| 50 mm (2") to 200 mm (8") $K =$ | $40 f_T$ | $120 f_T$ |
| 250 mm (10") to 350 mm (14") $K =$ | $30 f_T$ | $90 f_T$ |
| 400 mm (16") to 1200 mm (48") $K =$ | $20 f_T$ | $60 f_T$ |
| Minimum pipe velocity (mps) for full disc lift = | $100\sqrt{V}$ | $40\sqrt{V}$ |

Note. mps = metres per second

APPENDIX A

CRANE PHYSICAL PROPERTIES OF FLUIDS AND FLOW CHARACTERISTICS OF VALVES, FITTINGS AND PIPE A - 29

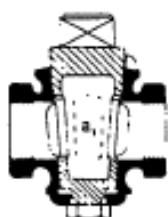
"K" FACTOR TABLE – SHEET 4 of 4

Representative Resistance Coefficients (K) for Valves and Fittings

(for formulas and friction data, see page A-26)
 ("K" is based on use of schedule pipe as listed on page 2-10)

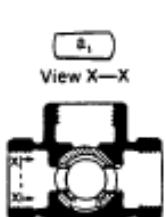
PLUG VALVES AND COCKS

Straight-Way

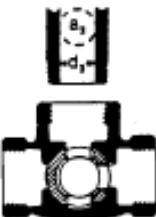


If: $\beta = 1$,
 $K_1 = 18 f_T$

3-Way



If: $\beta = 1$,
 $K_1 = 30 f_T$



If: $\beta = 1$,
 $K_1 = 90 f_T$

If: $\beta < 1$ $K_1 = \text{Formula 6}$

STANDARD ELBOWS

90°



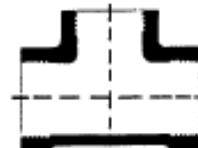
$K = 30 f_T$

45°



$K = 16 f_T$

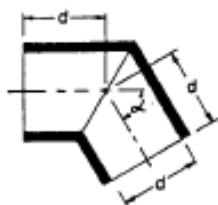
STANDARD TEES



Flow thru run $K = 20 f_T$

Flow thru branch $K = 60 f_T$

MITRE BENDS



| α | K |
|----------|----------|
| 0° | $2 f_T$ |
| 15° | $4 f_T$ |
| 30° | $8 f_T$ |
| 45° | $15 f_T$ |
| 60° | $25 f_T$ |
| 75° | $40 f_T$ |
| 90° | $60 f_T$ |

90° PIPE BENDS AND

FLANGED OR BUTT-WELDING 90° ELBOWS

$1.5 = r/d = 14 \text{ ft} = \text{Standard}$



| r/d | K | r/d | K |
|-------|----------|-------|----------|
| 1 | $20 f_T$ | 8 | $24 f_T$ |
| 1.5 | $14 f_T$ | 10 | $30 f_T$ |
| 2 | $12 f_T$ | 12 | $34 f_T$ |
| 3 | $12 f_T$ | 14 | $38 f_T$ |
| 4 | $14 f_T$ | 16 | $42 f_T$ |
| 6 | $17 f_T$ | 20 | $50 f_T$ |

The resistance coefficient, K_B , for pipe bends other than 90° may be determined as follows:

$$K_B = (n - 1) \left(0.25 \pi f_T \frac{r}{d} + 0.5 K \right) + K$$

n = number of 90° bends

K = resistance coefficient for one 90° bend (per table)

$$K_B = 31 \left(0.25 \pi \times 0.013 \times 3 + 0.5 \times 0.156 \right) + 0.156$$

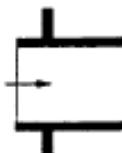
CLOSE PATTERN RETURN BENDS



$K = 50 f_T$

PIPE ENTRANCE

Inward Projecting

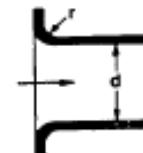


$K = 0.78$

| r/d | K |
|-----------|------|
| 0.00* | 0.5 |
| 0.02 | 0.28 |
| 0.04 | 0.24 |
| 0.06 | 0.15 |
| 0.10 | 0.09 |
| 0.15 & up | 0.04 |

*Sharp-edged

Flush



For K ,
see table

PIPE EXIT

Projecting



$K = 1.0$

Sharp-Edged



$K = 1.0$

Rounded



$K = 1.0$

"K" FACTOR TABLE – SHEET 1 of 4
Representative Resistance Coefficients (K) for Valves and Fittings
(“K” is based on use of schedule pipe as listed on page 2-10)

**PIPE FRICTION DATA FOR CLEAN COMMERCIAL STEEL PIPE
WITH FLOW IN ZONE OF COMPLETE TURBULENCE**

| Nominal Size | mm in. | 15 ½ | 20 ¾ | 25 1 | 32 1¼ | 40 1½ | 50 2 | 65, 80 2½, 3 | 100 4 | 125 5 | 150 6 | 200, 250 8, 10 | 300-400 12-16 | 450-600 18-24 |
|---------------------------|-----------|---------|---------|---------|----------|----------|---------|-----------------|----------|----------|----------|-------------------|------------------|------------------|
| Friction Factor (f_T) | | .027 | .025 | .023 | .022 | .021 | .019 | .018 | .017 | .016 | .015 | .014 | .013 | .012 |

NB MOODY FRICTION FACTOR

Solution:

Unit conversions

- $Q = 2300 \text{ l min}^{-1} = 0.0383 \text{ m}^3 \text{ s}^{-1}$
- $d = 128.2 \text{ mm} = 0.1282 \text{ m}$
- $\mu = 450 \text{ cP} = 0.45 \text{ Pa s}$

Total length $L = 85 \text{ m}$

Calculate the frictional pressure drop by first calculating v

$$v = \frac{Q}{A} = \frac{4 \cdot 0.0383}{\pi \cdot 0.1282^2} = 2.97 \text{ m s}^{-1} \quad (117)$$

We then calculate the Reynolds number to check which friction factor to use

$$Re = \frac{\rho v d}{\mu} = \frac{899 \cdot 2.97 \cdot 0.1282}{0.45} = 760.7 \quad (118)$$

$Re < 2000$ therefore we have laminar flow and the friction factor is given by

$$f = \frac{64}{Re} = \frac{64}{760.7} = 0.0841 \quad (119)$$

Calculate the pressure drop using K terms

$$\Delta p = K \frac{\rho v^2}{d} \quad (120)$$

Obtain the K values from the K -factor tables below using a β value of 1. The K values are

- Gate valve $\rightarrow K = 8 \cdot f_T$
- Weld elbow $\rightarrow K = 20 \cdot f_T$ (Wrong one is highlighted!)
- Angle valve $\rightarrow K = 150 \cdot f_T$
- Pipe $\rightarrow K = \frac{fL}{d} = \frac{0.0841 \cdot 85}{0.1282} = 55.76$

The total K value is

$$K_{total} = 8 \cdot 0.016 + 20 \cdot 0.016 + 150 \cdot 0.016 + 55.76 = 58.608 \quad (121)$$

The total frictional pressure drop is therefore

$$\Delta p = K \frac{\rho V^2}{d} = 58.608 \frac{899 \cdot 2.97^2}{0.1282} = 36.25 \text{ bar} \quad (122)$$

Next calculate the elevation pressure drop

$$\Delta p = \rho g \Delta z = 899 \cdot 9.81 \cdot 15 = 1.323 \text{ bar} \quad (123)$$

The total pressure drop is

$$\Delta p = 36.25 + 1.323 = 37.573 \text{ bar} \quad (124)$$

APPENDIX A
CRANE PHYSICAL PROPERTIES OF FLUIDS AND FLOW CHARACTERISTICS OF VALVES, FITTINGS AND PIPE A - 27

"K" FACTOR TABLE – SHEET 2 of 4

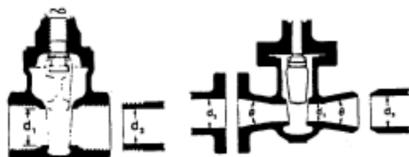
Representative Resistance Coefficients (K) for Valves and Fittings

(for formulas and friction data, see page A-26)

("K" is based on use of schedule pipe as listed on page 2-10)

GATE VALVES

Wedge Disc, Double Disc, or Plug Type



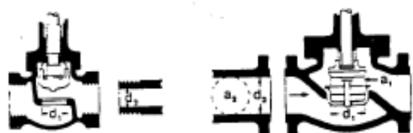
If: $\beta = 1, \theta = 0 \dots K_1 = 8 f_T$
 $\beta < 1 \text{ and } \theta < 45^\circ \dots K_1 = \text{Formula 5}$
 $\beta < 1 \text{ and } 45^\circ < \theta < 180^\circ \dots K_1 = \text{Formula 6}$

SWING CHECK VALVES



$K = 100 f_T$ $K = 50 f_T$
 Minimum pipe velocity (mps) for full disc lift
 $= 45\sqrt{V}$ $= 75\sqrt{V} \text{ except}$
 U/L listed $= 120\sqrt{V}$

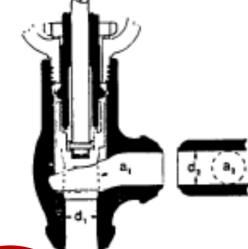
GLOBE AND ANGLE VALVES



If: $\beta = 1 \dots K_1 = 340 f_T$



If: $\beta = 1 \dots K_1 = 55 f_T$



If: $\beta = 1 \dots K_1 = 150 f_T$ If: $\beta = 1 \dots K_1 = 55 f_T$

All globe and angle valves,
whether reduced seat or throttled,

If: $\beta < 1 \dots K_2 = \text{Formula 7}$

LIFT CHECK VALVES



If: $\beta = 1 \dots K_1 = 600 f_T$
 $\beta < 1 \dots K_2 = \text{Formula 7}$
 Minimum pipe velocity (mps) for full disc lift
 $= 50\beta^2 \sqrt{V}$



If: $\beta = 1 \dots K_1 = 55 f_T$
 $\beta < 1 \dots K_2 = \text{Formula 7}$
 Minimum pipe velocity (mps) for full disc lift
 $= 170\beta^2 \sqrt{V}$

TLTING DISC CHECK VALVES



| Sizes | $\alpha = 5^\circ$ | $\alpha = 15^\circ$ |
|--|--------------------|---------------------|
| 50 mm (2") to 200 mm (8") | $40 f_T$ | $120 f_T$ |
| 250 mm (10") to 350 mm (14") | $30 f_T$ | $90 f_T$ |
| 400 mm (16") to 1200 mm (48") | $20 f_T$ | $60 f_T$ |
| Minimum pipe velocity (mps) for full disc lift = | $100\sqrt{V}$ | $40\sqrt{V}$ |

Note. mps = metres per second

APPENDIX A
CRANE PHYSICAL PROPERTIES OF FLUIDS AND FLOW CHARACTERISTICS OF VALVES, FITTINGS AND PIPE A - 29

"K" FACTOR TABLE – SHEET 4 of 4

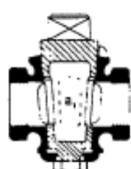
Representative Resistance Coefficients (K) for Valves and Fittings

(For formulas and friction data, see page A-26)

("K" is based on use of schedule pipe as listed on page 2-10)

PLUG VALVES AND COCKS

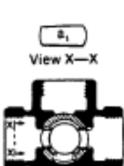
Straight-Way



$$\text{If: } \beta = 1, K_1 = 18 f_T$$

$$\text{If: } \beta < 1 \quad K_1 = \text{Formula 6}$$

3-Way



$$\text{If: } \beta = 1, K_1 = 30 f_T$$

STANDARD ELBOWS

90°



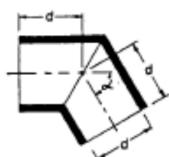
$$K = 30 f_T$$

45°



$$K = 16 f_T$$

MITRE BENDS



| α | K |
|----------|----------|
| 0° | $2 f_T$ |
| 15° | $4 f_T$ |
| 30° | $8 f_T$ |
| 45° | $15 f_T$ |
| 60° | $25 f_T$ |
| 75° | $40 f_T$ |
| 90° | $60 f_T$ |

90° PIPE BENDS AND FLANGED OR BUTT-WELDING 90° ELBOWS

$1.5 = r/d = 14 \text{ ft} = \text{Standard}$



| r/d | K | r/d | K |
|-----|----------|-----|----------|
| 1 | $20 f_T$ | 8 | $24 f_T$ |
| 1.5 | $14 f_T$ | 10 | $30 f_T$ |
| 2 | $12 f_T$ | 12 | $34 f_T$ |
| 3 | $12 f_T$ | 14 | $38 f_T$ |
| 4 | $14 f_T$ | 16 | $42 f_T$ |
| 6 | $17 f_T$ | 20 | $50 f_T$ |

The resistance coefficient, K_B , for pipe bends other than 90° may be determined as follows:

$$K_B = (n - 1) \left(0.25 \pi f_T \frac{r}{d} + 0.5 K \right) + K$$

n = number of 90° bends

K = resistance coefficient for one 90° bend (per table)

$$K_B = 31 \left(0.25 \pi \times 0.013 \times 3 + 0.5 \times 0.156 \right) + 0.156$$

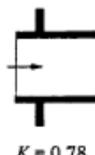
CLOSE PATTERN RETURN BENDS



$$K = 50 f_T$$

PIPE ENTRANCE

Inward Projecting

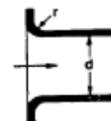


$$K = 0.78$$

| r/d | K |
|-----------|------|
| 0.00* | 0.5 |
| 0.02 | 0.28 |
| 0.04 | 0.24 |
| 0.06 | 0.15 |
| 0.10 | 0.09 |
| 0.15 & up | 0.04 |

*Sharp-edged

Flush



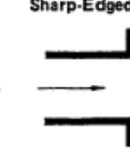
For K,
see table

PIPE EXIT



$$K = 1.0$$

Sharp-Edged



$$K = 1.0$$

Rounded



$$K = 1.0$$

[Question end]

Q.21

Question 21

Past exam question

- a) For a centrifugal compressor describe the condition known as surge.

[4 marks]

Solution:

Surge is when the compressor reaches its upper limits in pressure ratio and lower limit in flow rate. The compressor blades clear the working gas around themselves and spin free.

This prevents the compressors forming a back pressure to inhibit reverse flow. Reverse flow begins and causes an oscillation as it re-enters the compressor which again begins to provide back pressure. ✓

[4/4]

- b) Describe the concept of equivalent length as used in piping system pressure drop calculations. [2 marks]

Solution:

[2/2]

Equivalent length is used to calculate pressure drop arising from fittings and other flow features into an equivalent length of straight pipe pressure drop. ✓

- c) A polymer gel (density $\rho = 1000 \text{ kg m}^{-3}$ and viscosity $\mu = 0.01 \text{ Pa s}$) injection system is sketched below which has an inner diameter of 0.1 m. A volumetric flow of 10,000 barrel day $^{-1}$ is required. This is supplied to each well separately in sequence, so consider the two flow paths separately and at the full flowrate of 10,000 barrel day $^{-1}$. Calculate the pressure drops required to each well-head (B and C) assuming the pipe walls are smooth. Additional required information is available in Table 1 of the datasheet. [10 marks]

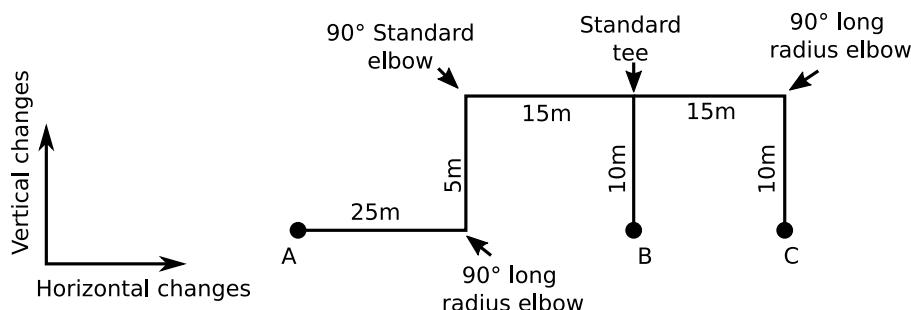


Figure 4: The flow diagram for the water injection system, and the pipeline. Please note the difference between vertical height changes and horizontal distance changes.

Solution:

The volumetric flow is

$$10000 \times 0.159/24/3600 = 0.0184 \text{ m}^3 \text{ s}^{-1}$$

[1/10]

✓ The pipe cross sectional area is

$$\pi \times 0.1^2/4 = 0.00785 \text{ m}^2$$

[1/10]

✓ Thus the average velocity is

$$0.0184/0.00785 = 2.343 \text{ m s}^{-1}$$

[1/10]

✓ The Reynolds number is then

$$Re = 1000 \times 2.343 \times 0.1/0.01 = 23431$$

[1/10]

✓

This is turbulent, thus the blasius correlation is appropriate.

$$f = 0.316 \times 23431^{-0.25} = 0.0255$$

[1/10]

✓

The pipeline lengths are 55 m (A→B) and 70 m (A→C). The connections for (A→B) are 110 L/D giving a total equivalent length of 66 m. The connections for (A→C) are 90 L/D giving a total equivalent length of 79 m.

[1/10]

The frictional pressure loss per length of pipe is as follows:

$$\frac{\Delta p}{L} = 1000 \times 0.0255 \times 2.343^2 / 2 / 0.1 = 701 \text{ Pa m}^{-1}$$

[1/10]

✓

[1/10]

This gives an overall frictional pressure loss of $701 \times 66 = 46300$ Pa (A→B) and $701 \times 79 = 55400$ Pa (A→C).

[1/10]

The overall height change for both lines is -5 m, thus the hydrostatic pressure loss is $1000 \times 9.81 \times (-5) = -49050$ Pa.

[1/10]

In total the pressure requirements are $46300 - 49050 = -2780$ Pa (A→B) and $55400 - 49050 = 6338$ Pa (A→C).

d) What equipment is required to maintain the required flowrate to each well? [4 marks]

Solution:

[4/4]

Expecting a sensible answer in-line with their results above. For the correct pressure values a pump is required at point A to add an additional 6338 Pa for injection into well C, but a control valve is required at point B to limit the flowrate as siphoning is occurring here. In practice both wells need control valves to control the flow into each well.

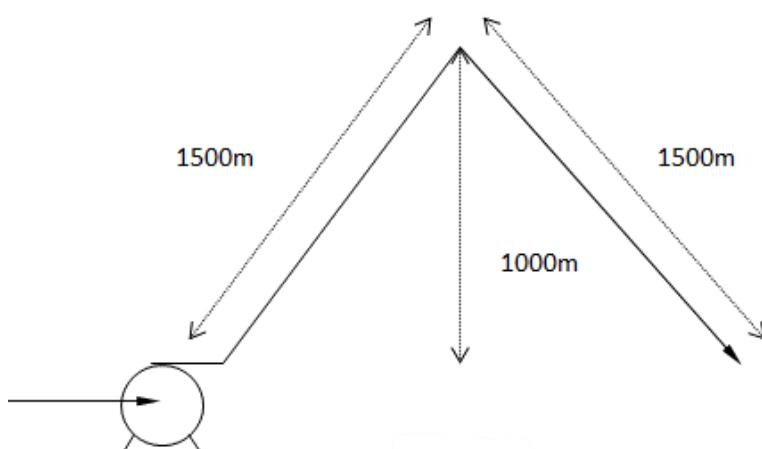
[Question total: 20 marks]

Q.22

Question 22

2: Single-phase fluid flow

The figure below shows a simplified route for an oil pipeline.



The oil has a specific gravity of 0.82. The inlet temperature to the pipeline is 30°C and it can be assumed that there is a 1°C temperature drop every 150 m. The True Vapour Pressure of the oil is 1000, 1200 and 1500 kPa at 10, 20 and 30°C respectively. At design conditions the total frictional pressure drop across the whole pipeline length is 500 kPa. You may ignore any pressure loss from fittings such as elbows or joints.

- a) Calculate the minimum pump discharge pressure to avoid vapour evolution from the oil. State any assumptions. [6 marks]

Solution:

Assume the oil is incompressible hence pressure drop due to friction will be linear. Thus 250 kPa pressure loss will occur in both rising and falling legs. The hydrostatic loss in the rising section will be;

$$\rho g h = 820 \times 9.81 \times 1000 / 1000 = 8044.2 \text{ kPa}$$

Hence the total pressure drop in the rising section will be

$$8044.2 \text{ kPa} + 250 = 8294.2 \text{ kPa}$$

[2/6]

At the apex of the pipeline the pipeline pressure must be above the vapour pressure of the oil to prevent vapour breakout. The temperature at the apex will be;

$$30 - 1500 \times 1/150 = 20^\circ C$$

[2/6]

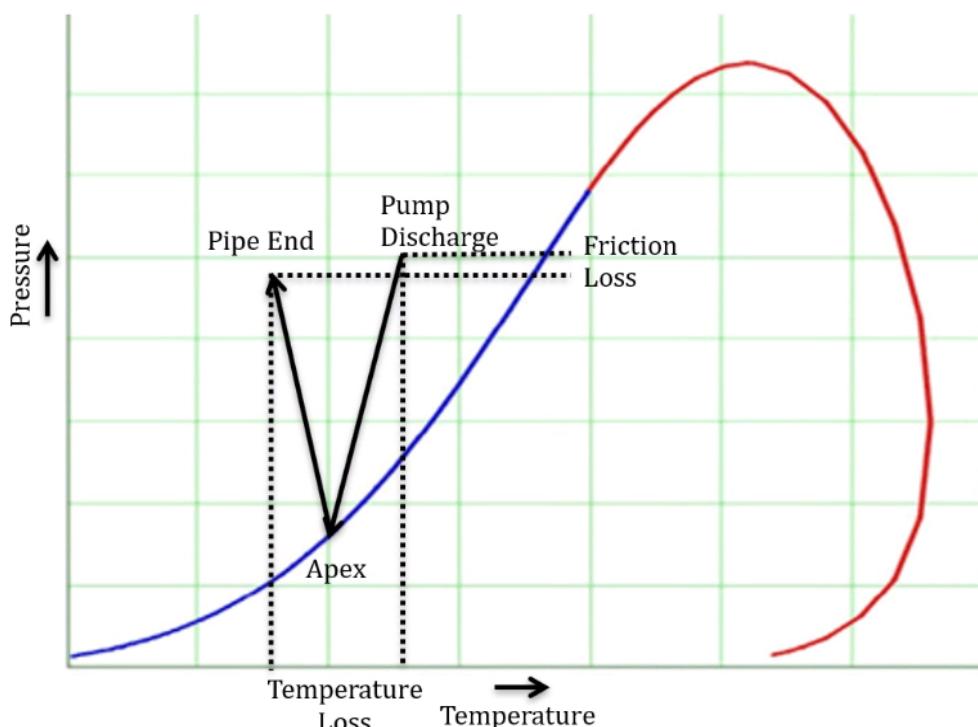
Hence vapour pressure is 1200kPa.

[2/6]

Thus minimum pump discharge pressure is $1200 + 8294 = 9494 \text{ kPa}$. After that hydrostatic pressure rises faster than frictional pressure loss thus the pipe is safe from vapour evolution.

- b) Sketch the pipeline pressure-temperature profile with respect to a typical oil phase envelope. Indicate the P,T conditions at the pipeline inlet, apex and pipeline outlet. [4 marks]

Solution:



- c) Explain the concept of equivalent length as used in the calculation of pressure drop in a piping system. [4 marks]

Solution:

Piping components - valves, bends, T-junctions etc. are represented by the length of a straight section of pipe which would result in the same pressure drop.

- d) The formula for single phase frictional pressure drop in a pipe is given by:

$$\Delta P_f = \frac{\rho f L v^2}{2 D}$$

where f is the Moody friction factor (proportional to $f \propto Re^{-0.25}$), L is the pipe length, v is the fluid velocity, D is the internal pipe diameter, ΔP_f is the frictional pressure loss, and ρ is the fluid density. An incompressible fluid is flowing in a horizontal section of pipe and the measured pressure drop is 2 bar.

- i) Calculate the pressure drop if the volumetric flow is halved. [3 marks]

Solution:

The Reynolds number will halve, thus f will increase by $2^{0.25} = 1.19$. Velocity will halve, hence the pressure drop will be $1.19 \times 0.5^2 \times 2 = 0.595$ bar. ✓

- ii) Calculate the pressure drop if the volumetric flow remains the same but the pipe diameter is halved instead. [3 marks]

Solution:

If the diameter is halved, the velocity will increase by a factor of 4. Re will increase by a factor of $4 \times 0.5 = 2$. Hence f will reduce by $0.5^{0.25} = 0.84$.

The new pressure drop becomes $2 \times 0.85 \times 4^2 / 0.5 = 53.76$ bar. ✓

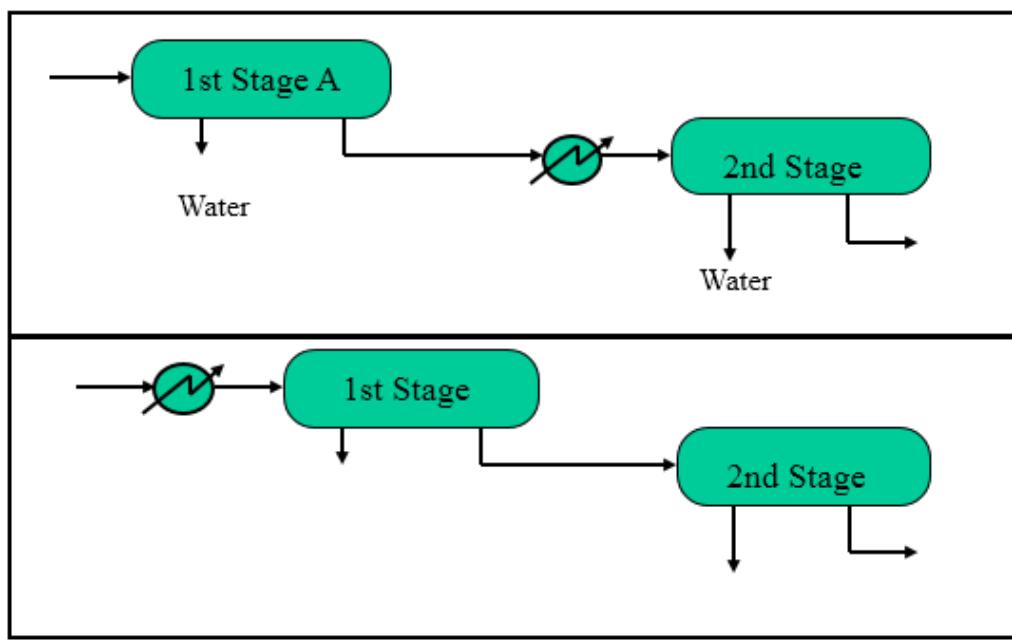
[Question total: 20 marks]

3 3: Separator

Q.23 Question 23

Prepare a conceptual design for an oil/gas/water separation train handling a relatively low API crude - 22 API. Previous experience of low API crudes in the North Sea indicates that oil-water separation can be difficult, especially if stable emulsions are formed. In order to understand some of the oil-water separation characteristics of this crude, you have commissioned a series of lab tests at varying water cuts and temperatures. The results are attached.

Consider the following arrangements. Identify the issues / benefits associated with the two options and select your preferred option.



Carry out basic separator sizing based on residence times inferred from the lab tests. Vessel weights are required for costing and dimensions are required for plant layout. Additionally, calculate the heat exchanger duty for the scheme. Identify requirement for an electrostatic coalescer and if required prepare a size and weight. Develop a control system and an equipment list for your selected arrangement.

Use the following guidelines and information:

- Typical L/D ratios for oil-water separation is ≈ 3 .
- Normal liquid level in separators is 50 %.
- Arrival temperature of fluids is 50 °C.
- Arrival pressure is 10 bara.
- Gas / oil ratio in first stage is $60 \text{ sm}^3 / \text{m}^3$ oil and the gas molecular weight is 20.
- Gas / oil ratio in first stage is $10 \text{ sm}^3 / \text{m}^3$ oil and the gas molecular weight is 35.
- 2nd stage separator operating pressure is to be determined to produce required oil TVP.

Operating and design pressure

The following is typical guidance used by process design companies.

Typical design pressures are as follows:

| Operating pressure barg | Design Pressure barg |
|-------------------------|----------------------|
| 0-10 | MOP + 1 bar |
| 10-35 | MOP + 10 % |
| 35-70 | MOP + 3.5 bar |
| > 70 | MOP + 5 % |

MOP = Maximum Operating Pressure

Design temperatures

Maximum design temperature = max. operating temperature + 15 °C

Minimum design temperature = min. operating temperature - 5 °C (or minimum ambient temperature)

Consideration for the min. design temperature must take into account any depressurisation that may occur during an emergency or shutdown situations.

Note that it is a conceptual design which means an accuracy of ±30 %.

Production data

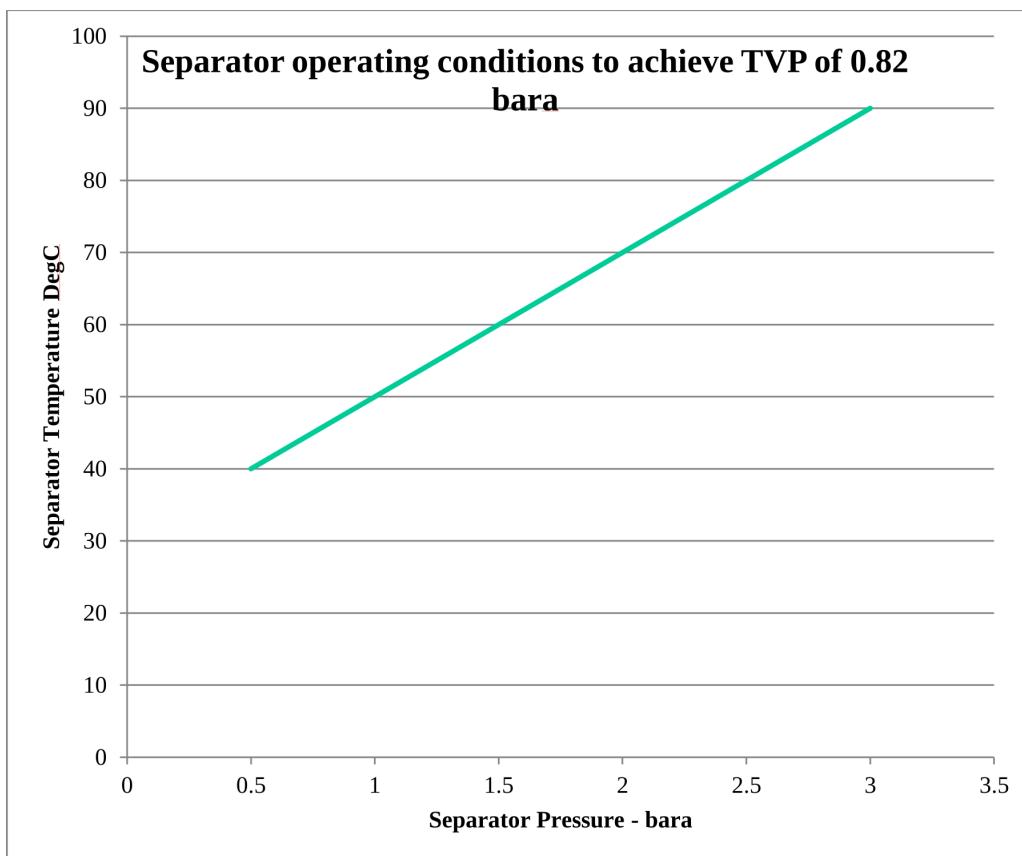
| | Flowrate bpd | Specific gravity | Specific heat capacity kJ kg ⁻¹ K ⁻¹ |
|-------|--------------|------------------|--|
| Oil | 89,000 | 0.83 | 2.1 |
| Water | 60,000 | 1.025 | 4.2 |

Oil specification

TVP = 0.82 bar at 37.8 °C

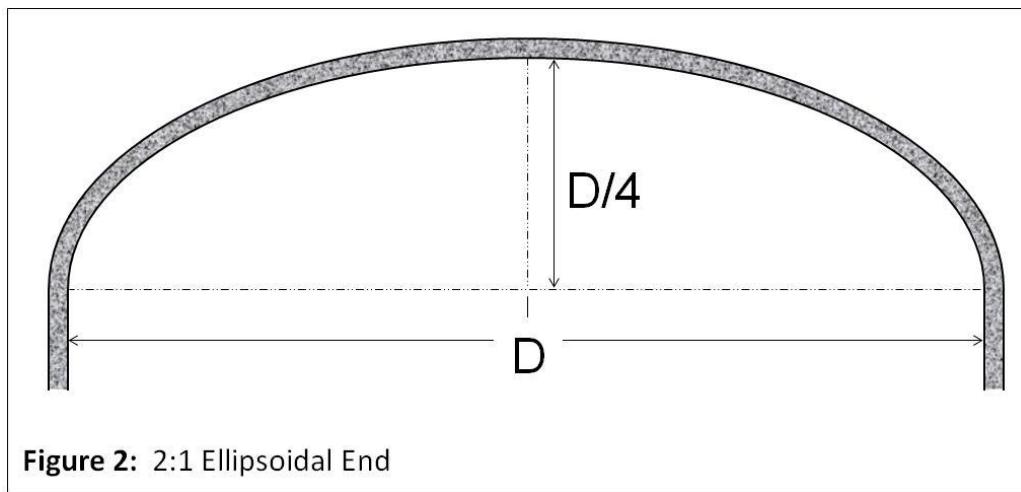
BS&W = 0.5 % v/v

TVP characteristics:



Volume of a vessel with a 2:1 ellipsoidal dished ends is given by the following expression:

$$V = \frac{\pi D^3}{12} + \frac{\pi D^2 L}{4} \quad (125)$$



Additional equations

The wall thickness is given by

$$t = \frac{PD}{20(SE - 0.06P)} \quad (126)$$

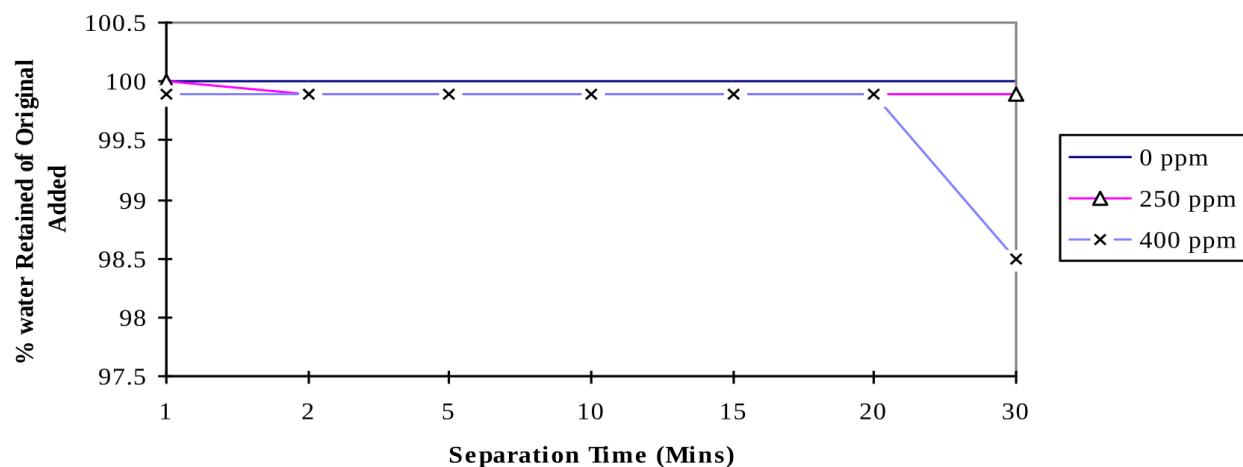
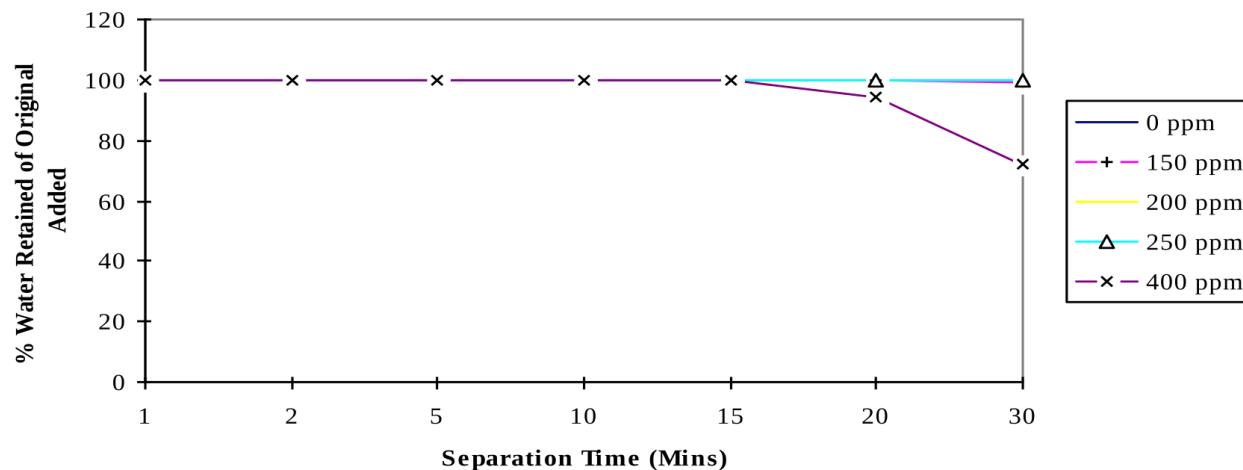
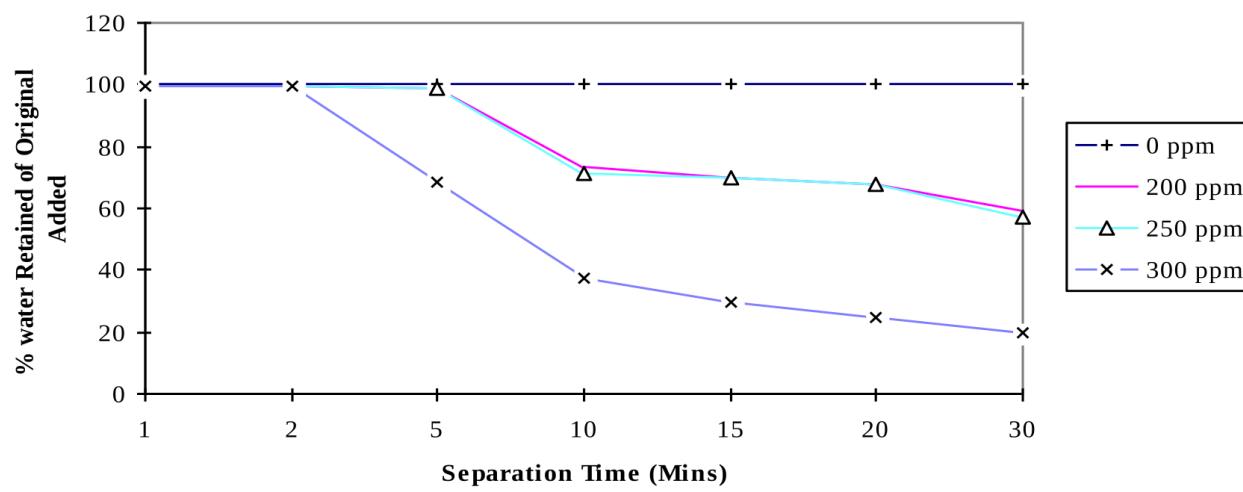
where t is the wall thickness in m, P is the design pressure in bara, S is the allowable stress in N mm⁻² (normally 155 N mm⁻² for carbon steel), E is the joint efficiency (assume $E = 1$), and D is the diameter in m.

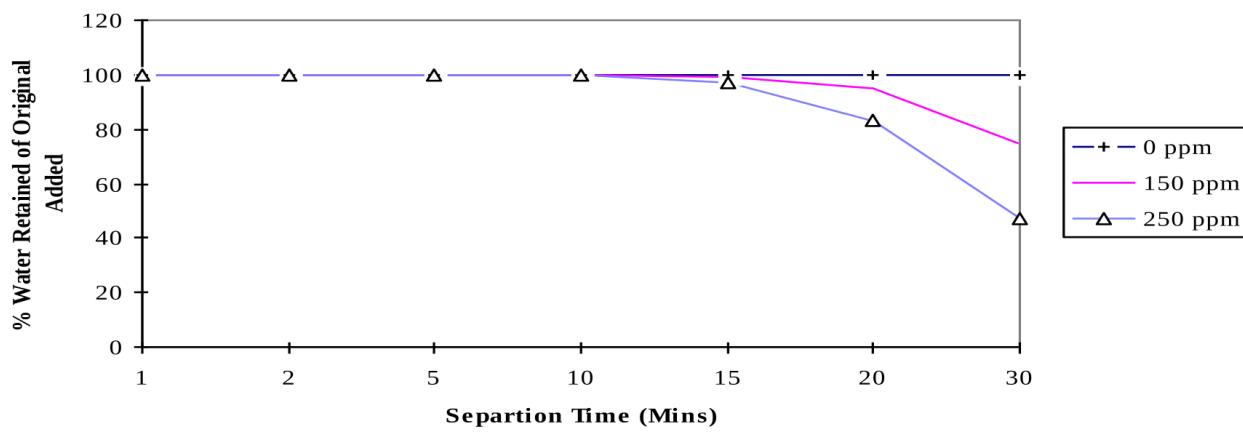
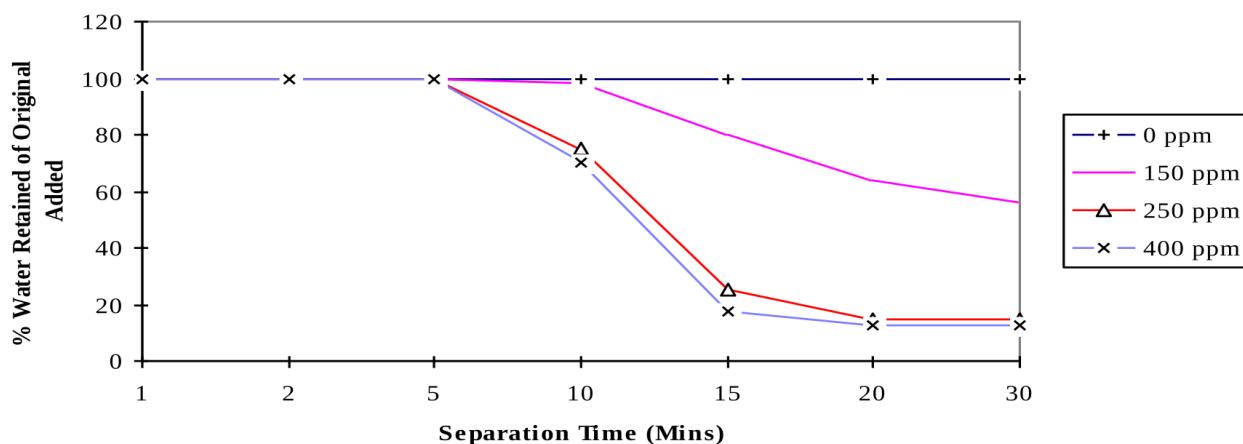
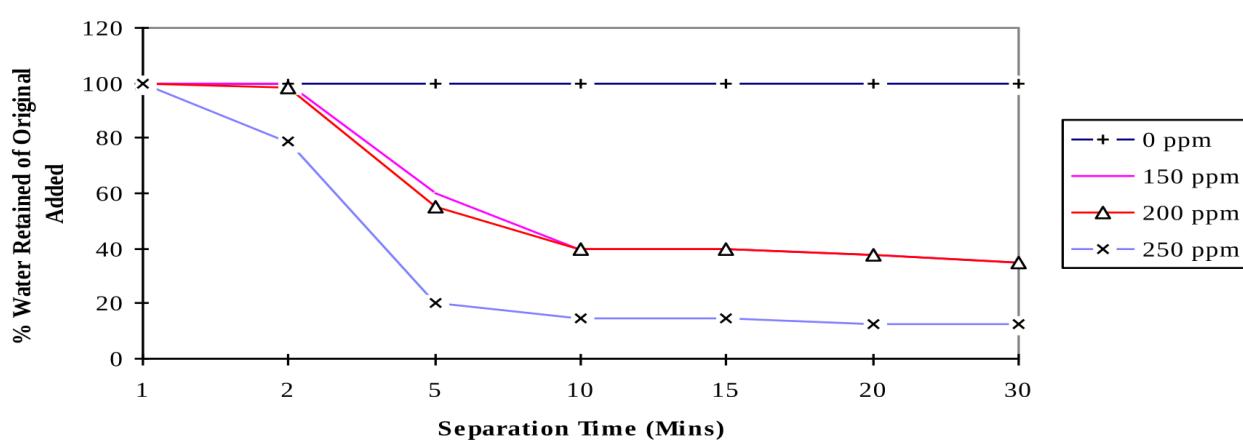
The surface area is given by

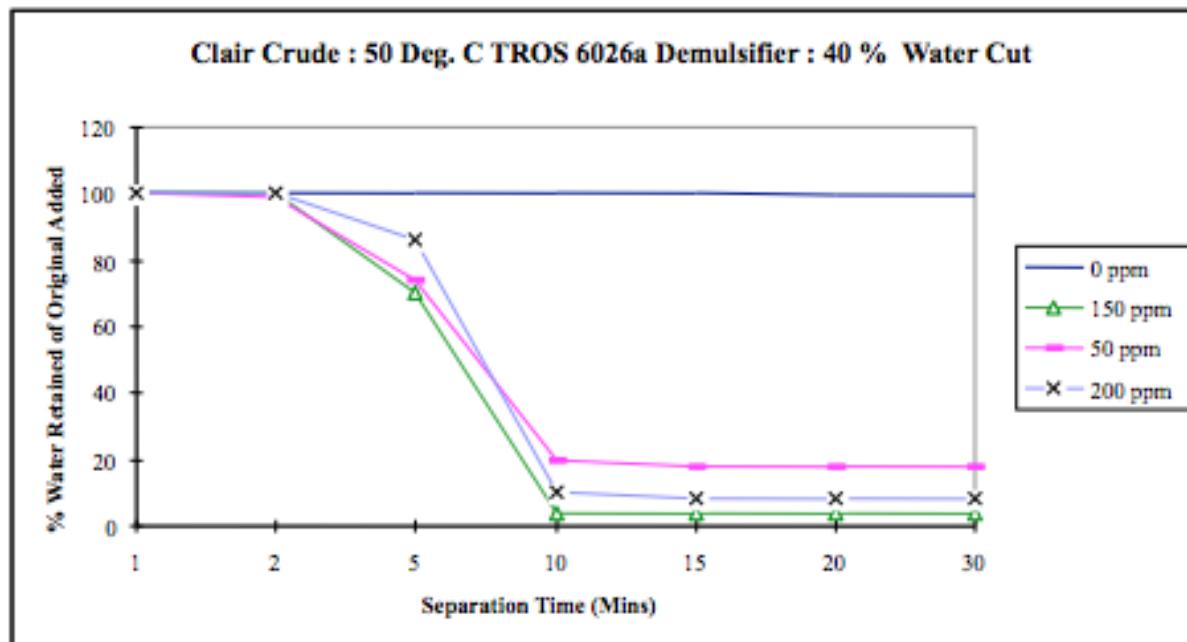
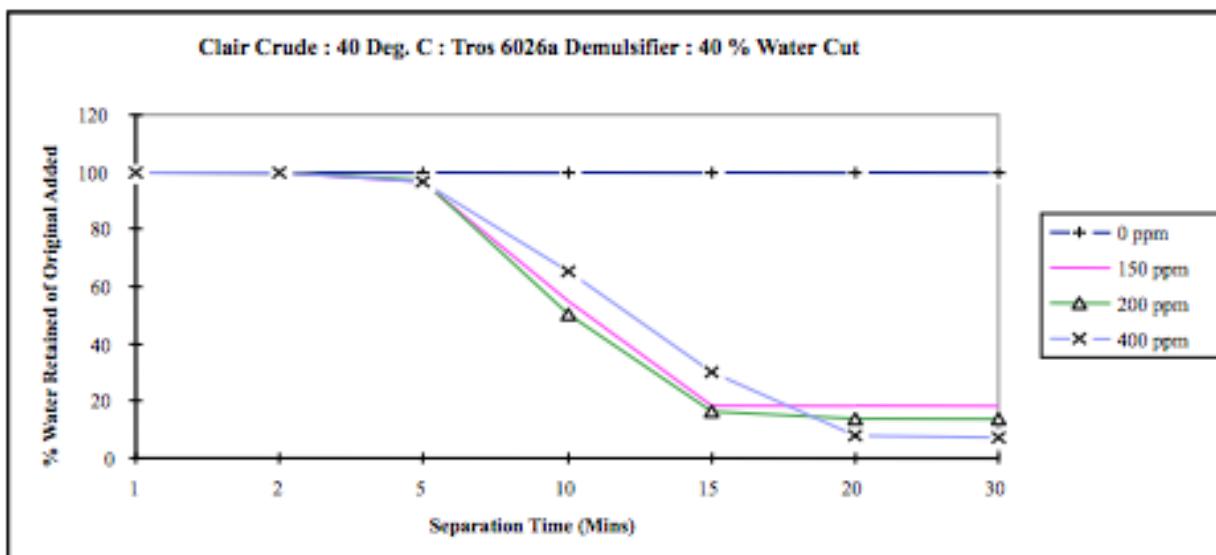
$$A = \pi DL + 0.8\pi D^2 \quad (127)$$

The vessel weight is given by

$$W(\text{kg}) = 8000At \quad (128)$$

Bottle tests**Clair Crude : 40 Deg. C : TROS 6026a Demulsifier : 10% Water Cut****Clair Crude : 50 Deg. C : TROS 6026a Demulsifier : 10 % Water Cut****Clair Crude : 90 Deg. C : TROS 6026a Demulsifier : 10% Water Cut**

Clair Crude : 40 Deg. C : TROS 6026a Demulsifier : 20% Water Cut**Clair Crude : 50 Deg. C : TROS 6026a Demulsifier : 20 % Water Cut****Clair Crude : 90 Deg. C : TROS 6026a Demulsifier : 20% Water Cut**



Equipment list

An equipment list will include a table showing:

- Equipment tag number e.g. V101
- A tag descriptor e.g. HP separator
- Number off - $1 \times 100\%$, $2 \times 50\%$ etc.
- Capacity per item
- Materials of construction
- Operating and design conditions

- Electrical load
- Size
- Dry weight
- Operating weight

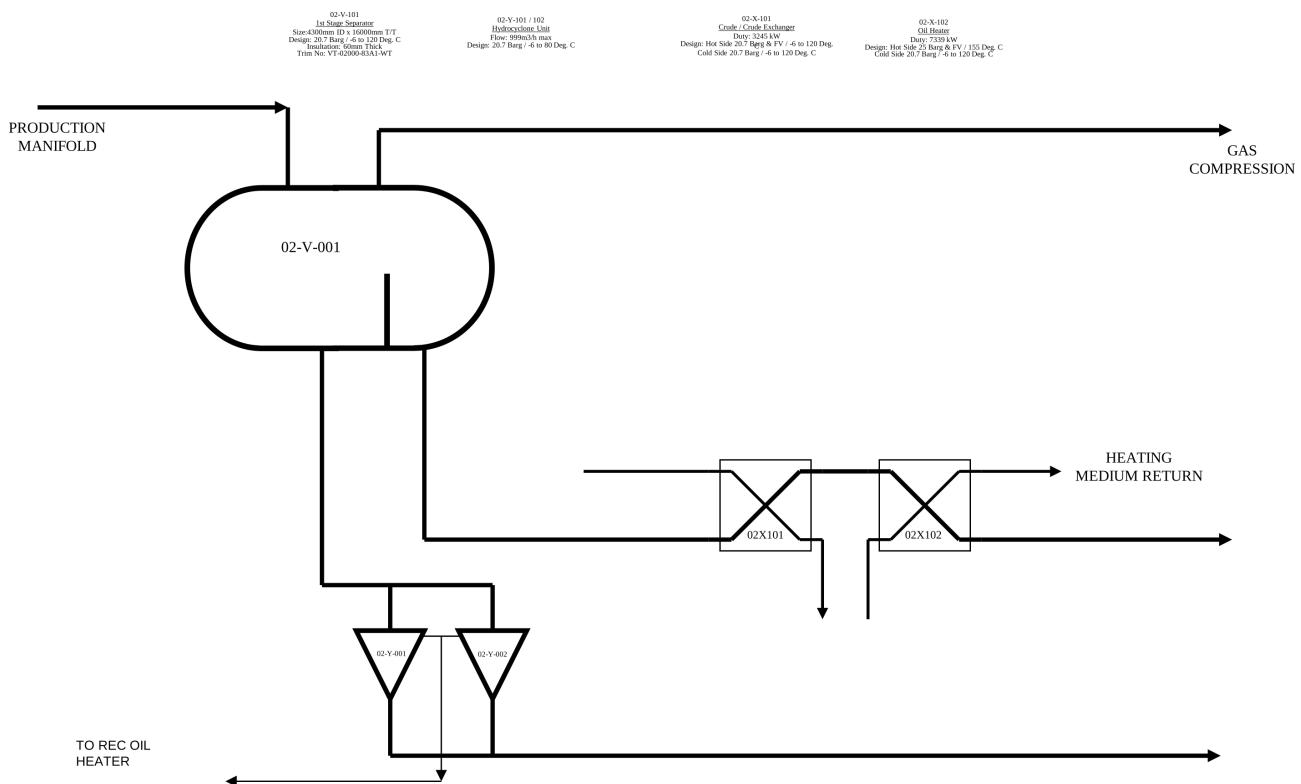
[Question end]

4 4: Produced water

Q.24

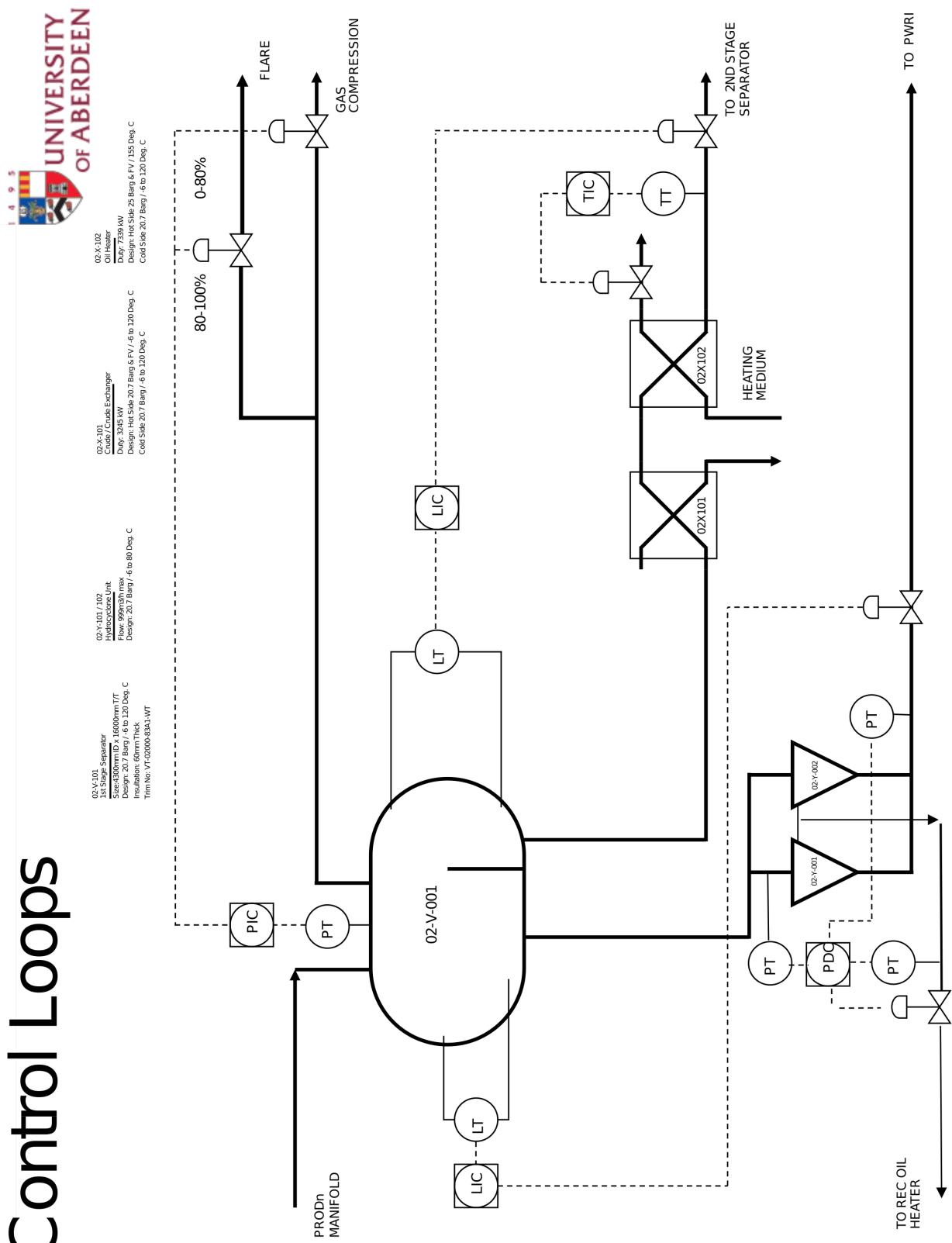
Question 24

Use the attached PFD of a bare three phase separator and associated hydrocyclone and develop this into a preliminary P&ID. Firstly consider the control system, then the alarm and trips, followed by the safety valves. Finally add in the pipework and valves required for maintenance.



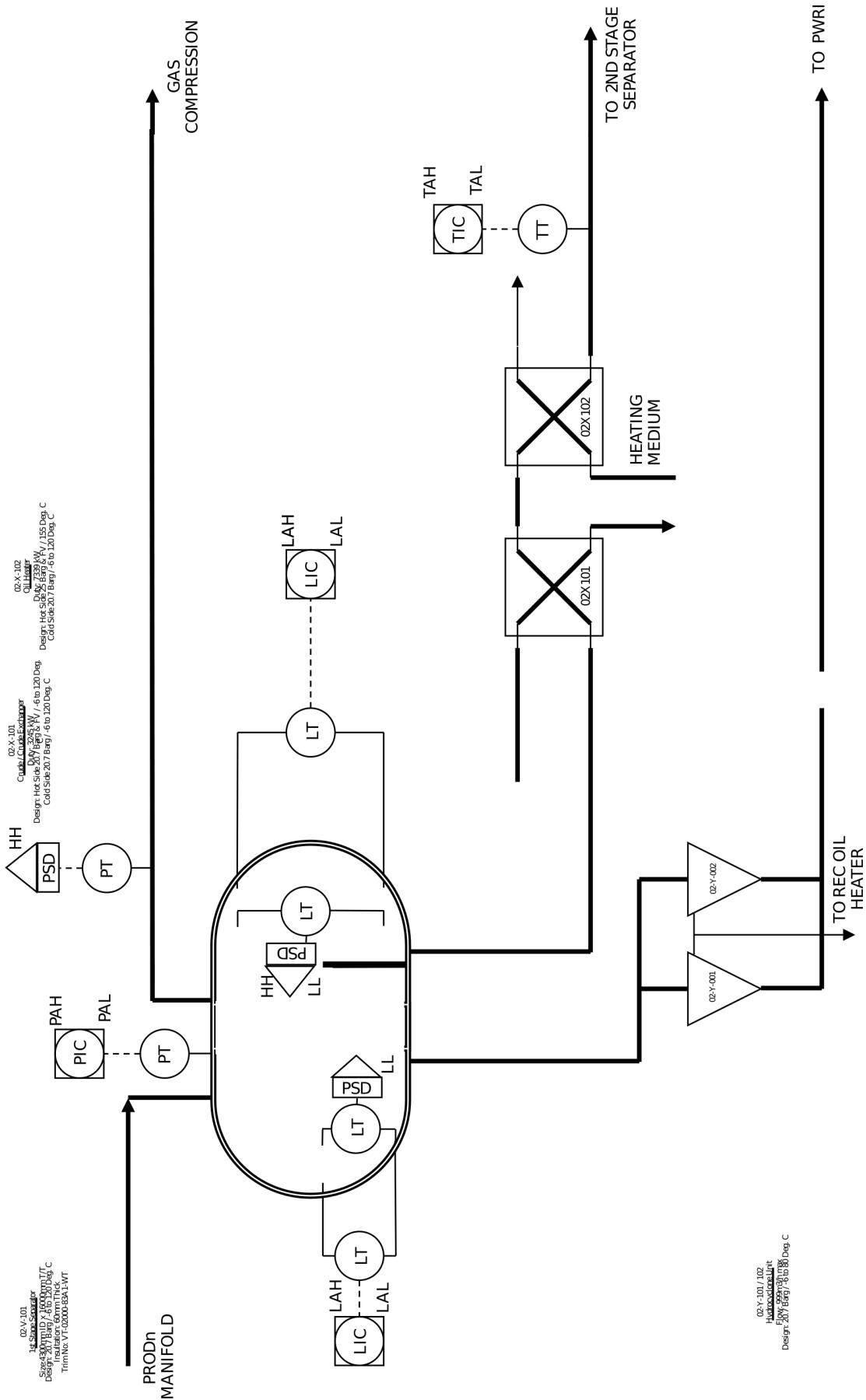
Solution:

Control Loops



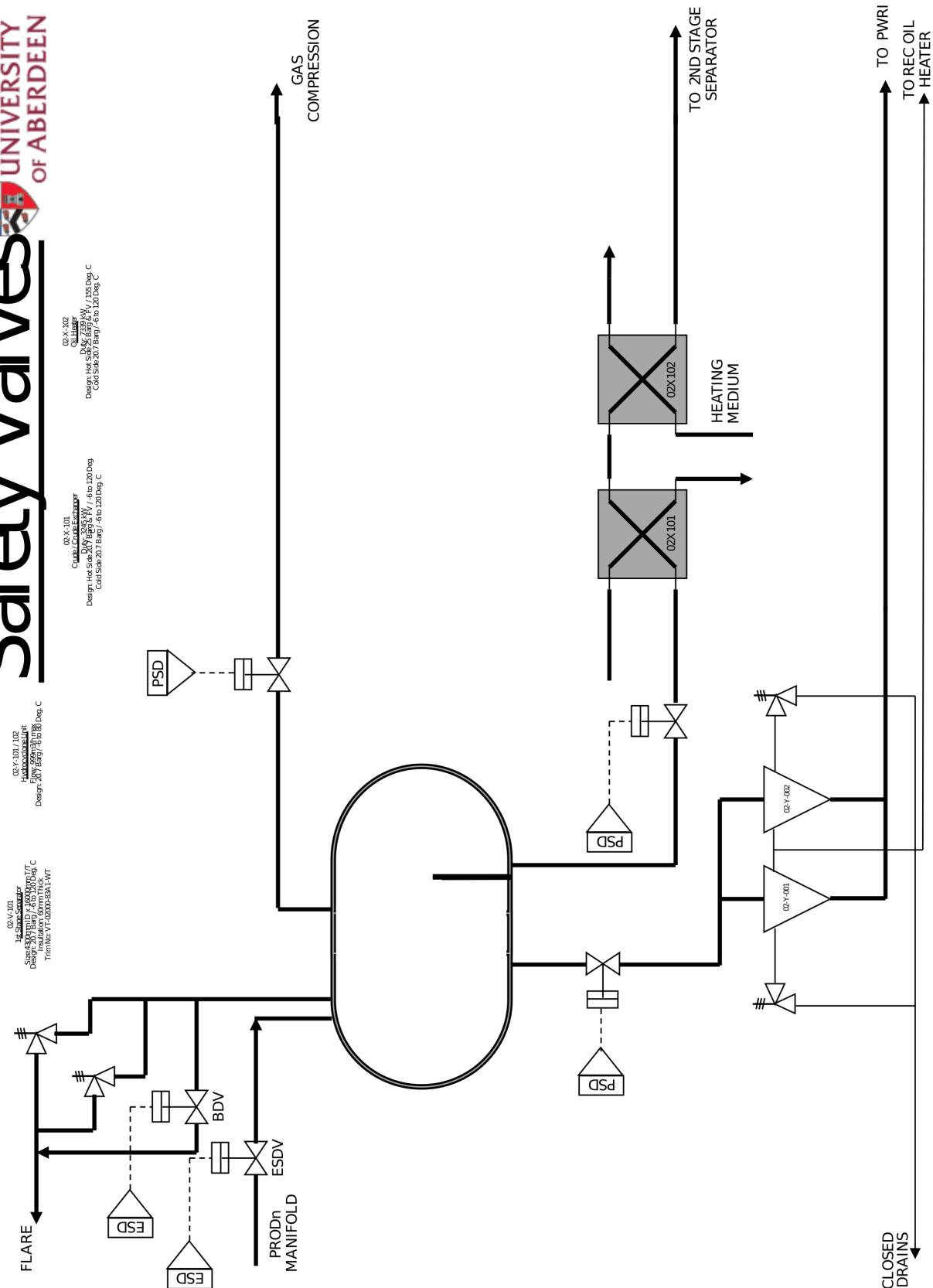


Alarm & Trip Devices



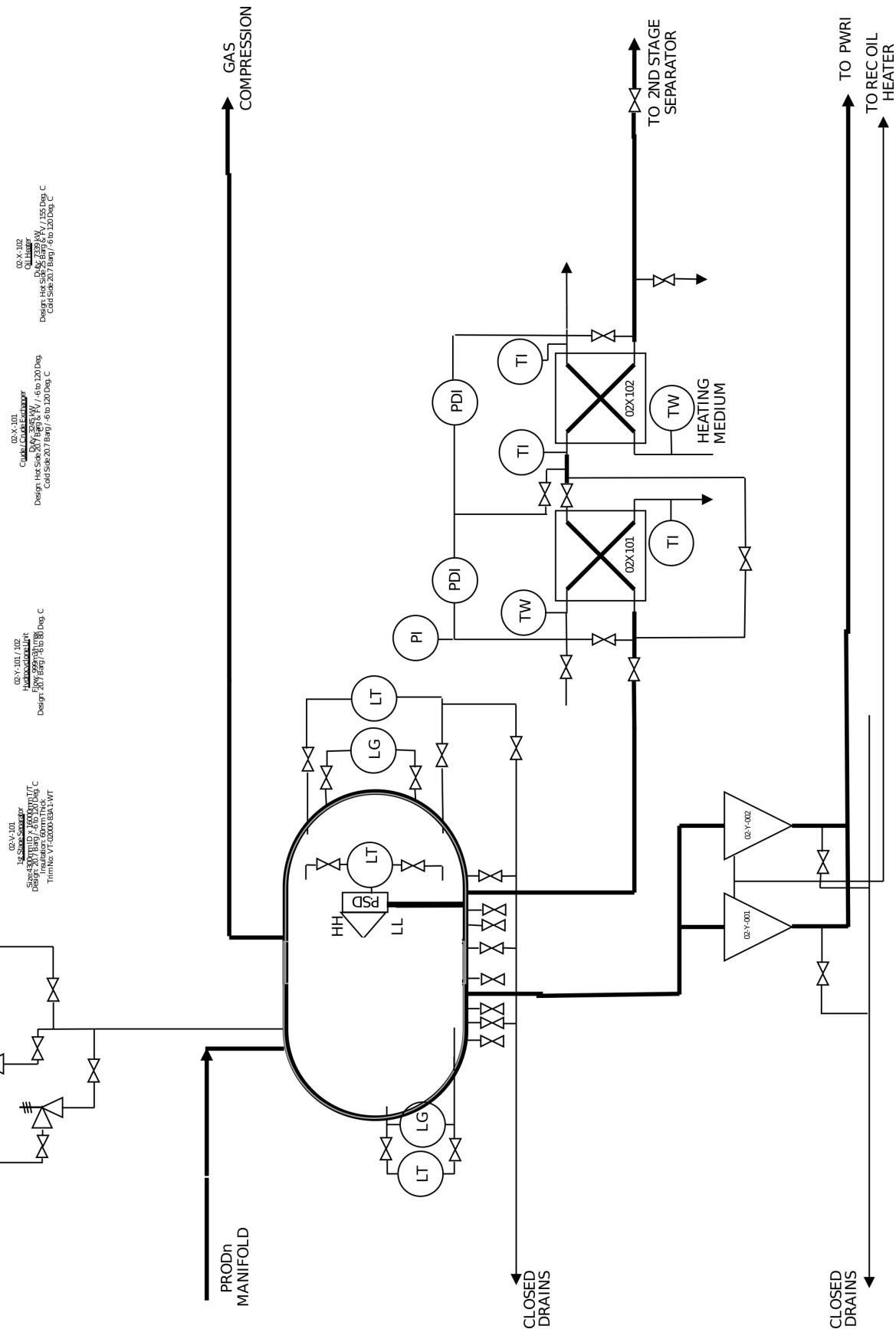


Safety Valves

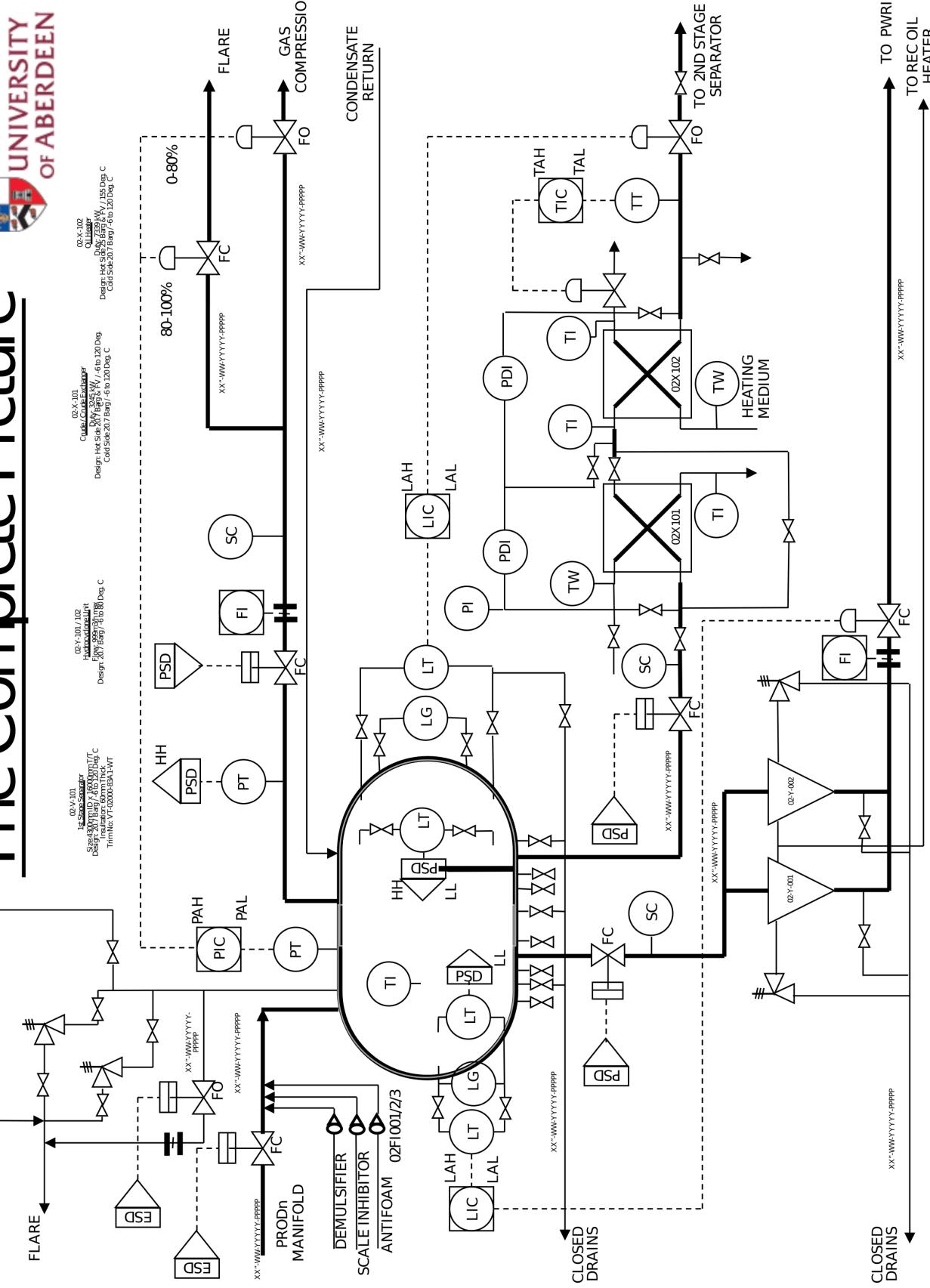


Maintenance Facilities

UNIVERSITY
OF ABERDEEN



The Complete Picture

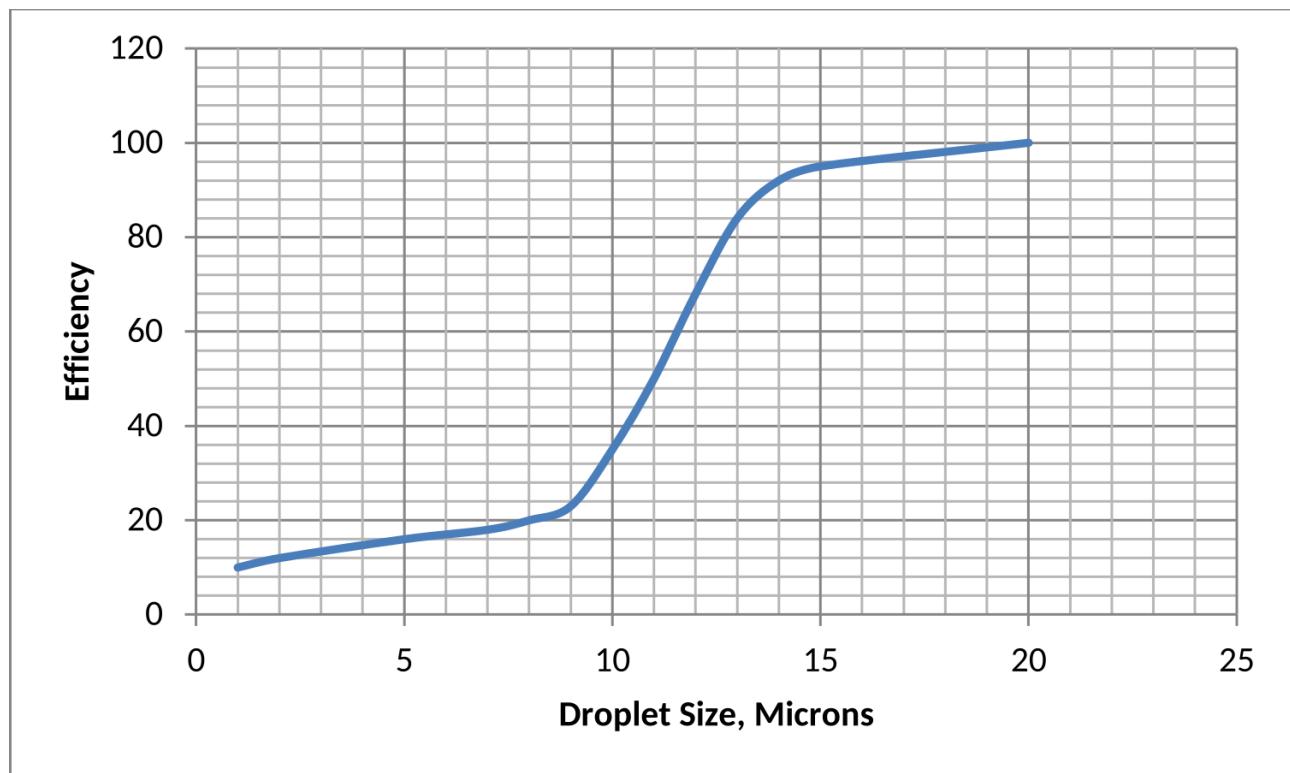


[Question end]

Q.25

Question 25

A hydrocyclone treating oil in water has the following oil removal characteristics.



The inlet oil concentration is 200 ppm by volume and the oil droplet size distribution entering the hydrocyclone is as follows:

| Oil droplet size (micron) | Volume % |
|---------------------------|----------|
| < 1 | 2 |
| 2 | 3 |
| 5 | 4 |
| 7 | 6 |
| 8 | 8 |
| 9 | 8 |
| 10 | 8 |
| 11 | 8 |
| 12 | 9 |
| 13 | 9 |
| 14 | 9 |
| 15 | 10 |
| > 20 | 16 |

- Calculate the predicted outlet oil in water concentration. Will this oil concentration be acceptable for overboard disposal?
- Repeat the calculation for the following distribution

| Oil droplet size micron | Volume % |
|-------------------------|----------|
| < 1 | 0.5 |
| 2 | 0.5 |
| 5 | 3 |
| 7 | 3 |
| 8 | 4 |
| 9 | 5 |
| 10 | 6 |
| 11 | 10 |
| 12 | 12 |
| 13 | 12 |
| 14 | 12 |
| 15 | 14 |
| > 20 | 18 |

Solution:

Use the provided separation efficiency vs droplet size graph in order to obtain the corresponding efficiency value for each quoted droplet size in the table provided. This is the percentage of that droplet size which will be removed from the water. We can then use this value in order to calculate the percentage value of oil which is retained in the water.

| Oil droplet size micron | Efficiency | Volume % | Oil % retained in water |
|-------------------------|------------|----------|-------------------------|
| < 1 | 10 | 2 | 1.8 |
| 2 | 12 | 3 | 2.64 |
| 5 | 16 | 4 | 3.36 |
| 7 | 18 | 6 | 4.92 |
| 8 | 20 | 8 | 6.4 |
| 9 | 23 | 8 | 6.16 |
| 10 | 35 | 8 | 5.2 |
| 11 | 50 | 8 | 4 |
| 12 | 68 | 9 | 2.88 |
| 13 | 84 | 9 | 1.44 |
| 14 | 92 | 9 | 0.72 |
| 15 | 95 | 10 | 0.5 |
| > 20 | 100 | 16 | 0 |
| | sum | 100 | 40.02 |

Now that we have calculated the % of oil retained in the water we can calculate the total outlet oil concentration

$$C_{outlet} = 200 \text{ ppm} \times 40.02 \% = 80.04 \text{ ppm} \quad (129)$$

The oil outlet concentration will be 80.04 ppm, specification for overboard disposal is 30 ppm. Hence this is unsuitable.

The second oil droplet distribution results in the following oil retention values:

| Oil droplet size micron | Efficiency | Volume % | Oil % retained in water |
|-------------------------|------------|----------|-------------------------|
| < 1 | 10 | 0.5 | 0.45 |
| 2 | 12 | 0.5 | 0.44 |
| 5 | 16 | 3 | 2.52 |
| 7 | 18 | 3 | 2.46 |
| 8 | 20 | 4 | 3.2 |
| 9 | 23 | 5 | 3.85 |
| 10 | 35 | 6 | 3.9 |
| 11 | 50 | 10 | 5 |
| 12 | 68 | 12 | 3.84 |
| 13 | 84 | 12 | 1.92 |
| 14 | 92 | 12 | 0.96 |
| 15 | 95 | 14 | 0.7 |
| > 20 | 100 | 18 | 0 |
| | sum | 100 | 29.24 |

$$C_{outlet} = 200 \text{ ppm} \times 29.24 \% = 58.48 \text{ ppm} \quad (130)$$

In this case the outlet oil distribution is 58.48 ppm which is still unsuitable.

[Question end]

Q.26 Question 26

Review the attached paper. Identify data which could be used to access the accuracy of suitable phase behaviour models. Use this information to investigate the accuracy of a range of phase behaviour models contained within ASPEN HYSYS, UNISIM etc. Identify an appropriate phase behaviour model.

Mutual Solubilities of Benzene and Water

Equilibria in the Two Phase Liquid-Liquid Region

WILLIAM H. THOMPSON and J. ROBERT SNYDER
The Pennsylvania State University, University Park, Pa.

The mutual solubilities in the liquid-liquid benzene-water system were studied from 100 to 460° F. at pressures of 1000 and 5000 p.s.i.g. The densities of the saturated liquid phases were also obtained. The phase-contacting apparatus consisted of an insulated high pressure cell with provisions for heating, pressurizing, sampling, and temperature and pressure measurements. The liquid-liquid system under study was contained in the cell by a moveable piston. Measurement of the piston motion provided data for calculation of density. To avoid emulsification problems, the equilibrium cell was not agitated, which resulted in the use of a lengthy equilibration period (24 hours). The solubility results are generally in good agreement with results that are available from other studies at different pressure and temperature conditions. The densities of the saturated phases are almost identical to pure solvent densities at low temperatures, but differ appreciably at high temperatures where the high solubilities have a strong influence on density.

ALTHOUGH a substantial amount of benzene-water mutual solubility data is available in the literature (1, 3-7, 10-13, 15, 16, 19-22), most of the data are limited to moderate temperatures and pressures, and many of the results are conflicting. This study (23) was undertaken to obtain liquid-liquid mutual solubility data at higher temperatures and pressures (up to 460° F. and 5000 p.s.i.g.), and to lay groundwork for subsequent solubility studies up to the critical solution conditions for the benzene-water system.

APPARATUS

The phase-contacting equipment consisted of a high-pressure cylinder fitted with a moveable piston for confining the benzene-water mixtures, and auxiliaries for sampling, pressure and thermal regulation, and for pressure, volume, and temperature measurement. The assembly is shown diagrammatically in Figure 1 while detailed information is available (23).

Nitrogen was hydraulically compressed in vessel I by fluid supplied from air-driven pump E. The pressure on the nitrogen was transmitted to distilled water (S) and in turn to the piston in equilibrium cell X. Water was employed for pressurizing cell X to avoid the possibility for trace contamination. Pressures were indicated on gages J and U and recorded via pressure transducer V. All gages were calibrated by dead-weight test gage and were accurate to about 25 p.s.i.

Since the equilibrium cell X was a smooth-bore cylinder, differential volumes could be calculated from measurements of the piston movement during sampling. Such measurements were gained from linear-variable-differential transformer Y (18) and machinists gage Z. The volume of a sample was measured to within 0.03 cc. The discharged sample was collected in chilled flask DD and weighed by analytical balance.

MATERIALS

The benzene used in this study was the pure grade (99 mole per cent minimum-purity) supplied by Phillips Petroleum Co. Several samples were analyzed by standard gas chromatographic procedures and found to be essentially

pure, since no other peaks appeared at maximum chromatograph sensitivity. Accordingly, the benzene was not further purified. The water was taken from the laboratory distilled water supply and redistilled in the absence of carbon dioxide in a simple batch still. Both materials were dispensed as required in the investigation from sealed vessels in which a nitrogen atmosphere was maintained.

PROCEDURE

For charging, the equilibrium cell was removed from the phase-contacting equipment, and both closures and the piston were removed. After cleaning the parts and inserting the piston, about 70-ml. each of water and benzene were withdrawn from storage and poured into the sample end of the cell so it was slightly overfilled and some liquid was forced out when the closure was screwed on. The other (pressurizing) end of the cell was filled with distilled water and closed. The charged cell was then installed in the phase-contacting equipment.

After filling surge cell S with distilled water and pressurizing the system to about 200 p.s.i. with nitrogen, the connections on tee W were slightly loosened to bleed out air and leave a continuous water couple between the water-surge cell S and the equilibrium-cell piston. Also, a portion of the charge was bled through the sampling valve to purge gases from the sampling lines.

With the equilibrium cell in the horizontal position, the system was pressurized by admitting nitrogen from cylinder K, and compressing the nitrogen by pumping oil into the system with pump E. During the heating period, the expansion of the cell charge was followed with the volume-measuring equipment to assure free piston motion. In some cases, nitrogen was vented from the system to relieve the pressure increase caused by the charge expansion.

After reaching the operating temperature and pressure, the cell normally remained in the horizontal position without agitation for a minimum of 24 hours. The cell assembly was then gently rotated into the vertical position and normally remained in this position for three hours to assure complete phase separation. With the sampling line and sample flask attached, valve AA was opened slightly and samples were drawn off at a rate less than 3-ml. per minute. The pressure on the piston was held constant

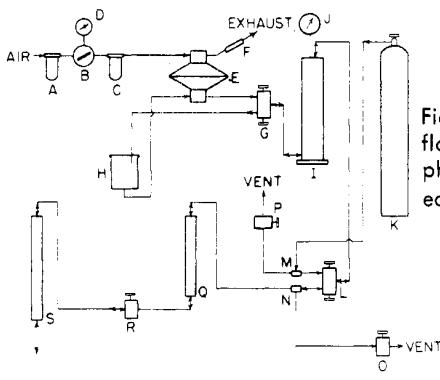


Figure 1. Simplified flow diagram of the phase-contacting equipment

Legend

- A = air filter
- B = air diaphragm valve
- C = air lubricator
- D = air pressure gage, 0–160 p.s.i.
- E = sprague pump
- F = muffler
- G = double valve for hydraulic fluid
- H = hydraulic fluid reservoir
- I = high pressure cell
- J = nitrogen pressure gage, 0–10,000 p.s.i.
- K = nitrogen cylinder
- L = double valve for nitrogen
- M, N = tee
- O, P = let-down valve
- Q = high pressure nitrogen surge cell
- R = valve
- S = high pressure water surge cell
- T = cross
- U = water pressure gage, 0–10,000 p.s.i.
- V = pressure transducer, 0–5,000 p.s.i.
- W = tee
- X = equilibrium cell
- Y = linear variable differential transformer, LVDT
- Z = vernier height gage
- AA = sampling valve
- BB = tee
- CC = thermocouple
- DD = sampling flask

by pump E. With the low sampling rate, enough cooling occurred in the sampling line that very little liquid was flash-vaporized in the Kjeldahl flask, and this small amount was effectively condensed in the bulb of the flask.

A sample of about 20-ml. was taken from both of the liquid phases in each experiment, and the upper, middle (interfacial), and lower portions of the charge were discarded. At the beginning and end of each sample, the vernier height gage reading was taken with the LVDT centered over its core. The difference between these values (the sample displacement) was converted with a calibration factor to give the volume that the sample occupied under the operating conditions.

After the cell charge had been completely withdrawn, valve AA was opened with full pressure on the piston to assure that the seal between the piston and the cell was intact.

ANALYSIS

The benzene-rich phase was analyzed for water by the Fischer method (8) using a potentiometric end-point (2). The sample, which had become heterogeneous after reducing the temperature from the higher equilibrium

value, was made homogeneous by adding a weighed amount of standard water-in-methanol solution. A weighed portion of the liquid was analyzed, and the solubility was calculated with an allowance for the water added in the methanol standard solution.

The water-rich phase was analyzed for benzene by slowly vaporizing the sample into a purified air stream, passing the vapors over copper oxide at 1450–1500° F., absorbing the resulting carbon dioxide in a known volume of standard sodium hydroxide solution, and determining the amount of sodium hydroxide that was not used to react with the carbon dioxide. The determination of unreacted sodium hydroxide was made by titrating with standard hydrochloric acid after precipitating the carbonates with excess barium chloride.

For both phases, the filled and empty sample flasks were weighed on an analytical balance to give the total sample weight by difference. Phase densities were then calculated using the sample volumes determined earlier.

Calibration of Cell Volume. The volume of the cell was calibrated by charging the cell with pure degassed water, and drawing off samples at various temperatures and pressures. The piston displacement for each sample was obtained by the difference between the height gage readings. After smoothing the experimental results to even temperature, the corresponding volumes were calculated from the weights of the water samples using specific volume data from Keenan and Keyes (14). In this way, conversion factors (cc. per linear inch of cell) for use in determining sample volumes were obtained that are functions of both temperature and pressure.

Equilibration. This work differs from almost all previous high-temperature solubility studies in that no agitation was used to help establish equilibrium rapidly. Here, agitation was avoided to preclude emulsion formation which could become a problem in subsequent work at temperatures nearer the critical solution temperature. In this study, the solute diffused through a solvent layer that remained stagnant after the initial heating period was completed. However, during the equilibration period, the cell was horizontal to provide a maximum interfacial area and minimum diffusive distance.

The approach to equilibrium in 24 hours for the dilute solutions was calculated for 100° F. using Fick's law and estimated (24) diffusivities. The boundary conditions used in the diffusion equation were: the solvent is uniformly saturated with solute after an infinitely long contact period; the solvent is always saturated with solute at the interface; the initial solute concentration in the solvent is zero; and the concentration gradient is always zero at the cell wall furthest from the interface. At 100° F., the diffusivity of benzene in water was taken as 6×10^{-5} ft.² per hr; and that of water in benzene, 3×10^{-4} . A rectangular cross-section was chosen to simplify the calculations. It was found that the approach to equilibrium in 24 hours was more than 99.9 per cent for the water diffusing into the benzene, and about 92 per cent for the benzene in water. This predicted approach to equilibrium for benzene in water would be unacceptable if it did not represent the limiting minimum approach. However, in practice, such factors as thermal agitation during heating and enhanced diffusivities at elevated temperatures result in considerably higher rates of mass transfer.

A series of experiments was performed at 100° F. which verified that equilibrium was established in 24 hours. The results for benzene dissolved in water are illustrated in Figure 2, and a similar pattern was found for water dissolved in benzene.

EXPERIMENTAL RESULTS

The experimental solubility and density results of this study are listed in Table I, and the smoothed results in

Table II. As indicated by asterisks, some data were discarded. Most of these poor results accumulated in the early developmental stages of the study due to poor sampling equipment and techniques. After reducing the sampling line volume and decreasing the sampling rate, both phases were successfully sampled without entraining a portion of the other liquid phase.

The solubility data are plotted on Figures 3 and 4. The water solubility line passes through all the experimental points for 1000 p.s.i.g., as does the benzene solubility curve at 5000 p.s.i.g. Although the points for water solubility at 5000 p.s.i.g. are scattered slightly, a line is well-defined. At 400° F., the line begins curving sharply as the critical solution temperature is approached. The worst data were obtained for the water phase at 1000 p.s.i.g. Since most of the 340 and 400° F. points were known to be high as a result of difficulties with irregularly high analytical blanks, the curve was drawn below them. The 5000 p.s.i.g. behavior was influential when drawing the 1000 p.s.i.g. curve.

Compared with solubility, the density data are scattered. However, although the results appear worse on the expanded scales of Figures 5 and 6, the average deviation of the points from the curve is less than one per cent of the density. In view of the secondary nature of these data, the reproducibility is considered satisfactory.

COMPARISON OF SOLUBILITY RESULTS

Both liquid phases of the benzene-water system have been extensively studied at low temperatures by other investigators, and results at higher temperatures are available for the water phase. The results of this study are compared with the literature data in Figures 3 and 4.

In the benzene phase, the logarithm of solubility was found to be linear with reciprocal temperature which is typical for the solubility of water in hydrocarbons. An increase in pressure results in a uniform decrease in solubility such that the 5000 p.s.i.g. isobar is below and about parallel to the 1000 p.s.i.g. isobar. By extrapolation, the atmospheric pressure isobar should be parallel to this work, but at higher solubilities. This is in agreement with the results of Stavely, Jeffes, and Moy (20). The solubilities of Groschuff (10) and Hill (12) are also higher and almost parallel. Although the other data that are shown in Figure 3 do not correspond exactly, the over-all agreement of the results of this study with those of others is considered excellent for the benzene phase.

Typical temperature behavior for hydrocarbons was also observed in the water phase, where a 4000-p.s.i. pressure increase was found to cause a large increase in solubility. A proportionate change should be observed between this 1000-p.s.i.g. work and the low pressure work of others, but the change is very small. Nevertheless, if the two

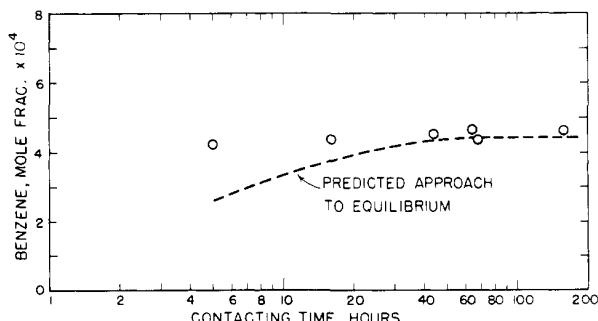


Figure 2. Benzene dissolved in water at 100° F. and 1000 p.s.i.g. as a function of contacting time

Table I. Experimental Solubility and Density Results

| Temp., °F. | Benzene Phase | | Water Phase | |
|---------------|--|--------------------|--|--------------------|
| | Water concn., mole frac. × 10 ⁴ | Density, g./cc. | Benzene concn., mole frac. × 10 ⁴ | Density, g./cc. |
| 1000 P.S.I.G. | | | | |
| 100.3 | 48.7 | 0.865 | 4.34 | 0.993 |
| 100.1 | 46.6 | 0.861 | 4.54 | 0.992 |
| 99.9 | 41.6 | 0.856 | 4.33 | 0.994 |
| 99.9 | 41.8 | 0.858 | 4.61 | 0.991 |
| 101.5 | 43.3 | 0.859 | ... | ... |
| 101.0 | 42.1 | 0.856 | 4.60 | 0.992 |
| 160.3 | 124 | 0.816 | 6.63 | 0.975 |
| 219.0 | 274 | 0.779 | 8.56 | 0.952 |
| 218.0 | ... | ... | 9.84 | 0.941 |
| 280.0 | 558. | 0.761 | * | 0.913 |
| 280.3 | ... | ... | 14.3 | 0.920 |
| 281.0 | ... | ... | 18.6 | 0.922 |
| 340.0 | 1000 | 0.673 | 44.2† | 0.859 |
| 339.3 | * | 0.686 | 42.7† | 0.905 |
| 339.3 | ... | 0.680 | 41.6† | 0.854 |
| 340.3 | ... | ... | 46.5† | 0.879 |
| 341.0 | ... | ... | 36.0† | 0.878 |
| 397.3 | 1770 | 0.641 | ... | ... |
| 396.3 | ... | ... | 85.3 | 0.815 |
| 399.7 | * | 0.575 | 81.0 | 0.830 |
| 398.0 | ... | 0.582 | 94.2† | 0.838 |
| 399.0 | ... | ... | 72.6 | 0.837 |
| 5000 P.S.I.G. | | | | |
| 100.3 | 38.1 | 0.885 | 5.61 | 1.005 |
| 161.0 | 105 | 0.848 | * | 0.988 |
| 159.7 | * | 0.851 | 8.95 | 0.987 |
| 160.3 | 110 | 0.850 | ... | ... |
| 219.7 | * | 0.812 | 14.1 | 0.966 |
| 220.7 | 263 | 0.820 | ... | 0.961 |
| 279.3 | 505 | 0.781 | 22.8 | 0.922 |
| 339.3 | 959 | 0.753 | 43.1 | 0.897 |
| 400.7 | 1530 | 0.713 | 98.3 | 0.867 |
| 460.7 | 4320 | 0.660 | 265 | 0.787 |
| 459.7 | 4600 | 0.664 | 248 | 0.790 |

* = No analysis was performed. † = Discarded data. ‡ = Results known to be high due to a high analytical blank.

100° F. points from this study were very slightly higher, the agreement with other low-temperature data would be excellent.

The high-temperature solubility results of Jaeger (13) required conversions and corrections for comparison with the solubilities from this study. In his pioneer work, Jaeger analyzed by simply measuring the volumes which separated after cooling the samples to room temperature. Accordingly room-temperature density data were used to convert the solubility results from the reported volumetric units. In addition, since Jaeger incorrectly assumed that the solubility of benzene in water at room temperature was negligible, 4×10^{-4} mole fraction units were added to each experimental solubility.

Since the pressure for the work of Jaeger increased with temperature, the results should converge on this work until intersection occurs with the 1000-p.s.i.g. curve at about 460° F. This behavior is not observed. Jaeger's results are lower than the results of this work for all temperatures. However, Jaeger's results for the xylenes are almost identically low when compared with the data of Pryor and Jentoft (17). Similar departures are observed in comparing the toluene data with that of Bohon and Claussen (6). Apparently, the 10-minute equilibration period used by Jaeger did not produce an equilibrium system.

The two isobars that were plotted in Figure 4 from the smoothed results of Kudchadker and McKetta (15) do not agree with this work or the work of others. These two isobars (for 60 and 450 p.s.i.a) indicate very unusual tem-

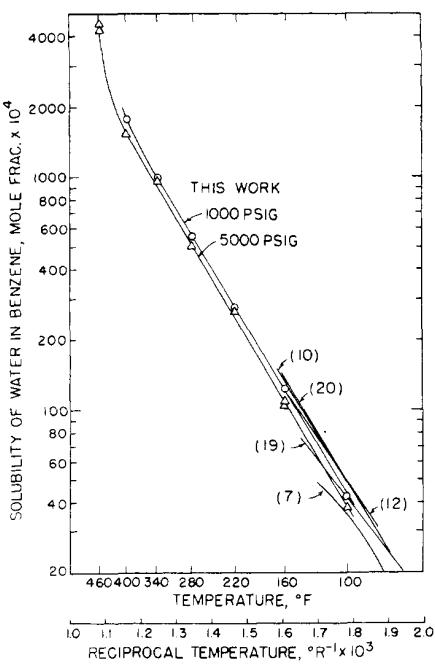


Figure 3. Solubility of water in benzene
The following work of others was performed at one atmosphere pressure or less: Clifford (7), Groschuff (10), Hill (12), Rosenbaum and Walton (19), and Stavely, Jeffes, and Moy (20).

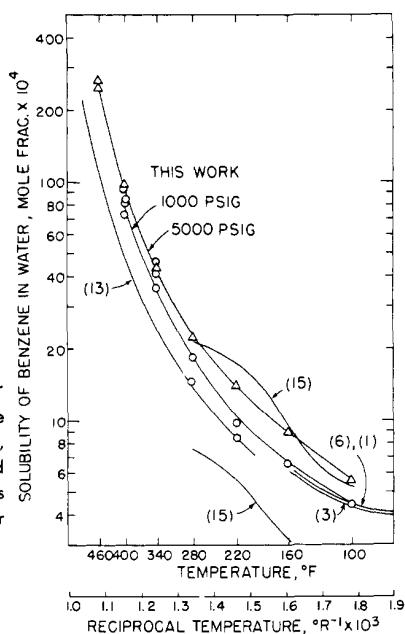


Figure 4. Solubility of benzene in water
The following work of others was performed at the vapor pressure of the system: Alexander (1), Arnold et al. (3), Bohon and Claussen (6), and Jaeger (13). Two curves are shown for the results of Kudchadker and McKetta (15), the upper for 450 p.s.i.a., and the lower, 60 p.s.i.a.

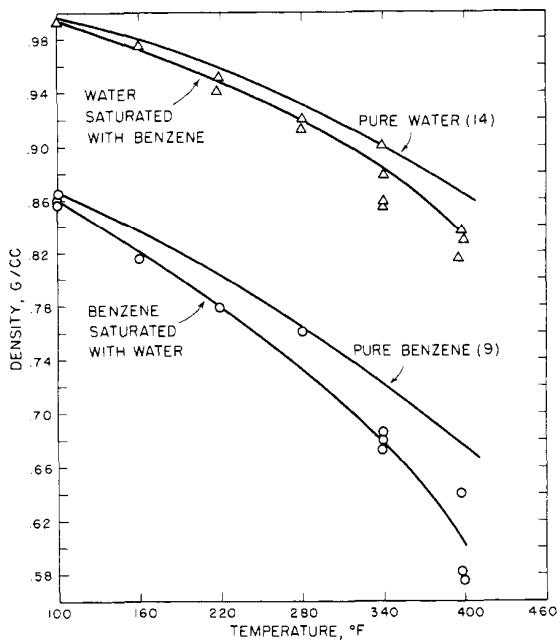


Figure 5. Density of the saturated liquid phases at 1000 p.s.i.g.

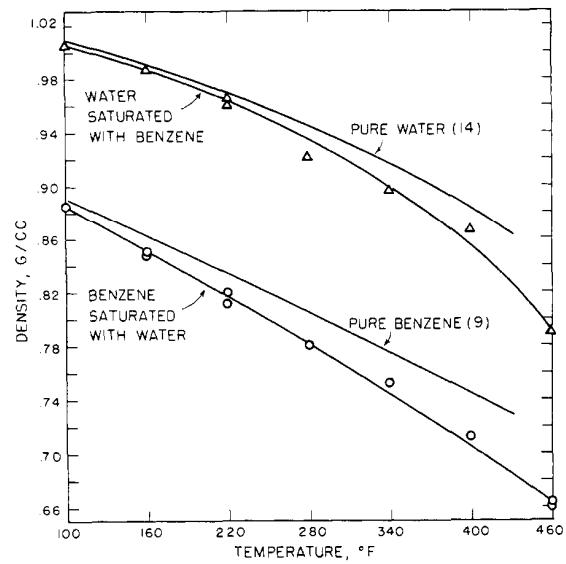


Figure 6. Density of the saturated liquid phases at 5000 p.s.i.g.

Table II. Experimental Solubility and Density Results

| Temp., °F. | Benzene Phase | | | Water Phase | | | | |
|---------------|------------------------------|-----------------------------|-------------|-----------------|------------------------------|-----------------------------|-------------|-----------------|
| | Mole frac. × 10 ⁴ | Wt. frac. × 10 ⁴ | Moles/liter | Density, g./cc. | Mole frac. × 10 ⁴ | Wt. frac. × 10 ⁴ | Moles/liter | Density, g./cc. |
| 1000 P.S.I.G. | | | | | | | | |
| 100 | 42.3 | 9.78 | 0.0467 | 0.860 | 4.46 | 19.3 | 0.0246 | 0.993 |
| 160 | 121 | 28.2 | 0.129 | 0.821 | 6.65 | 28.8 | 0.0360 | 0.972 |
| 220 | 274 | 64.5 | 0.279 | 0.779 | 10.1 | 43.6 | 0.0530 | 0.948 |
| 280 | 558 | 134 | 0.544 | 0.732 | 18.6 | 80.1 | 0.0942 | 0.919 |
| 340 | 1000 | 250 | 0.944 | 0.679 | 36.0 | 154 | 0.174 | 0.881 |
| 400 | 1780 | 475 | 1.58 | 0.601 | 82.0 | 346 | 0.369 | 0.831 |
| 5000 P.S.I.G. | | | | | | | | |
| 100 | 38.1 | 8.80 | 0.0432 | 0.885 | 5.61 | 24.3 | 0.312 | 1.005 |
| 160 | 106 | 24.6 | 0.116 | 0.851 | 9.05 | 39.2 | 0.0495 | 0.987 |
| 220 | 245 | 57.5 | 0.261 | 0.817 | 14.1 | 60.8 | 0.0751 | 0.965 |
| 280 | 500 | 120 | 0.520 | 0.781 | 22.8 | 98.0 | 0.117 | 0.935 |
| 340 | 905 | 224 | 0.927 | 0.745 | 43.0 | 184 | 0.211 | 0.899 |
| 400 | 1620 | 445 | 1.74 | 0.706 | 98.0 | 411 | 0.449 | 0.855 |
| 460 | 4500 | 1590 | 5.83 | 0.663 | 255 | 1020 | 1.03 | 0.791 |

perature behavior as well as a very strong influence of pressure on solubility. The smoothed atmospheric pressure results were so low that they could not be plotted on Figure 4.

The density results are compared with data for the pure solvents in Figures 5 and 6. As shown, at low temperatures, the small amount of dissolved solute causes a small decrease in solvent density. As temperature increases, solubility increases, and the density departure becomes larger.

ACKNOWLEDGMENT

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LITERATURE CITED

- (1) Alexander, D.M., *J. Phys. Chem.* **63**, 1021 (1959).
- (2) Almy, E.G., Griffin, W.C., Wilcox, C.S., *Ind. Eng. Chem. Anal. Ed.* **12**, 392 (1940).
- (3) Arnold, D.S., Plank, C.A., Erickson, E.E., *CHEM. ENG. DATA SER.* **3**, 253 (1958).
- (4) Berkengheim, T.I., *Zavodsk. Lab.* **10**, 592 (1941); *C.A.* **40**, 6961-8.
- (5) Black, C., Joris, G.G., Taylor, H.S., *J. Chem. Phys.* **16**, 537 (1948).
- (6) Bohon, R.L., Claussen, W.F., *J. Am. Chem. Soc.* **73**, 1571 (1951).
- (7) Clifford, C.W., *J. Ind. Eng. Chem.* **13**, 631 (1921).
- (8) Fischer, K., *Angew. Chem.* **48**, 394 (1935).
- (9) Glanville, J.W., Sage, B.H., *Ind. Eng. Chem.* **41**, 1272 (1949).
- (10) Groschuff, E., *Z. Elektrochem.* **17**, 348 (1911).
- (11) Gross, P.M., Saylor, J.H., *J. Am. Chem. Soc.* **53**, 1744 (1931).
- (12) Hill, A.E., *Ibid.* **45**, 1143 (1923).
- (13) Jaeger, A., *Brennstoff-chem.* **4**, 259 (1923).
- (14) Kennan, J.H., Keyes, F.G., "Thermodynamic Properties of Steam," John Wiley, New York, 1936.
- (15) Kudchadker, A.P., McKetta, J.J., *Hydrocarbon Process Petrol. Refiner* **41** (3), 191 (1962).
- (16) Niimi, A., *Suimen Kemistilehti* **11A**, 19 (1938); *C.A.* **32**, 4861.
- (17) Pryor, W.A., Jentoft, R.E., *J. CHEM. ENG. DATA* **6**, 36 (1961).
- (18) Reitzel, J., Simon, J., Wollser, J.A., *Rev. Sci. Inst.* **28**, 828 (1957).
- (19) Rosenbaum, C.K., Walton, J.H., *J. Am. Chem. Soc.* **52**, 3568 (1930).
- (20) Stavely, L.A.K., Jeffes, J.H.E., Moy, J.A.E., *Trans. Faraday Soc.* **39**, 5 (1943).
- (21) Tarasenkov, D.N., Polozhintzeva, E.N., *Ber.* **65B**, 184 (1932); *C.A.* **26**, 2363.
- (22) Tarasenkov, D.N., Polozhintzeva, E.N., *Zh. Obshch. Khim.*, Khim. Ser. I, 71 (1931); *C.A.* **25**, 4762.
- (23) Thompson, W.H., M.S. thesis, The Pennsylvania State University, University Park, Pa., 1962.
- (24) Treybal, R.E., "Mass-Transfer Operations," p. 108, McGraw-Hill, New York, 1955.

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Kinematic Viscosity of Liquid Rubidium from 67° to 688° C.

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The kinematic viscosity of pure rubidium was measured from 67° to 688° C. in an all-metal capillary viscometer. The calibration of the apparatus without the use of a reference fluid is described, and experimental results are compared with low-temperature data for rubidium and with data for potassium and sodium. The correlated kinematic viscosity of rubidium, in centistokes, may be expressed in terms of temperature, in degrees Kelvin, as: $\log (\mu/\rho) = 250/T - 1.15$.

THE ANTICIPATED USE of alkali metals as working fluids, heat transfer media, and lubricants in high-temperature energy-conversion systems has spurred considerable research on the thermophysical properties of these fluids. However, there are still many areas in which reliable data are meager or lacking. Experimental measurement of liquid alkali-metal viscosity is in progress in several laboratories, and new experimental data have been reported recently for potassium (17). The viscosities of cesium, potassium, sodium, and lithium have been measured at temperatures as high as 1150° C. (for potassium) by damped torsional oscillation methods (1, 2, 8, 9, 17), and that of sodium has been determined to 600° C. by capillary viscometry (4, 5). On the other hand, the experimental viscosity of rubidium has not been reported for temperatures above 220° C.

As part of an Air Force sponsored research program on the influence of contaminants in liquid alkali-metal systems, an all-metal capillary viscometer has been developed in this laboratory. Extensive viscosity measurements have been made on pure rubidium from 67° to 688° C., and it is the purpose of this paper to present the resulting experimental kinematic viscosity data.

EXPERIMENTAL APPARATUS AND MATERIALS

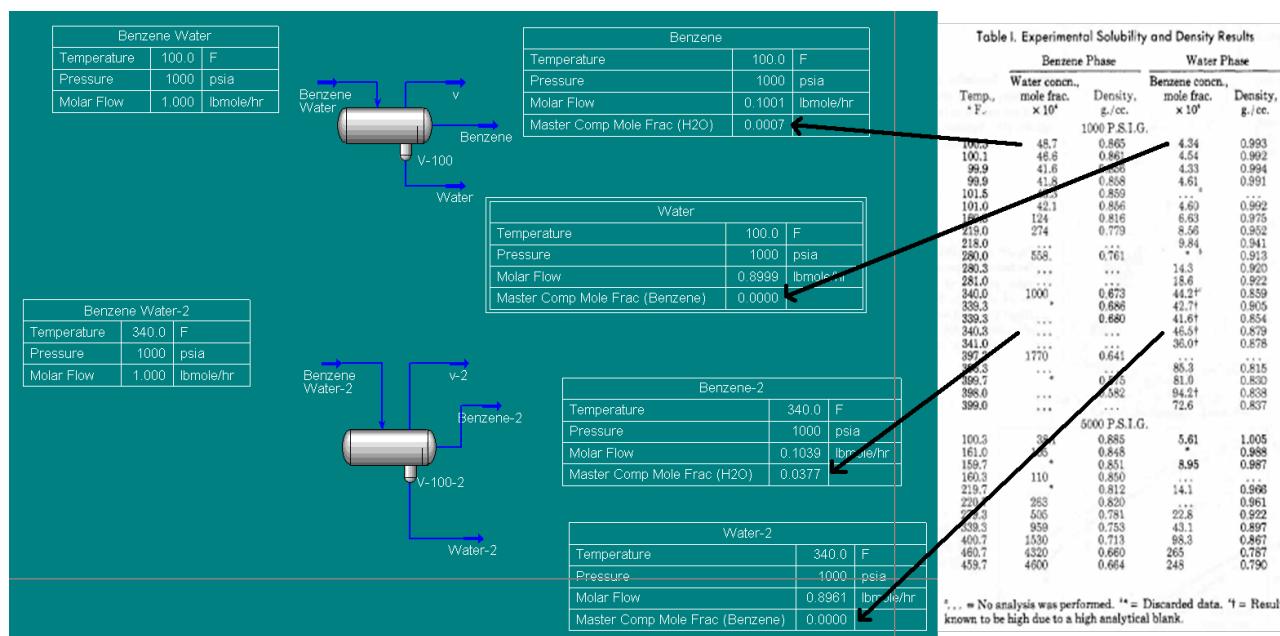
Rubidium Samples. Two batches of pure rubidium were obtained commercially for use in this program. Cesium represented the major impurity in each batch, being present to the extent of 0.4 and 0.02 wt. per cent, respectively. The rubidium was analyzed for trace metal constituents by emission spectroscopy; for alkali metals

Solution:

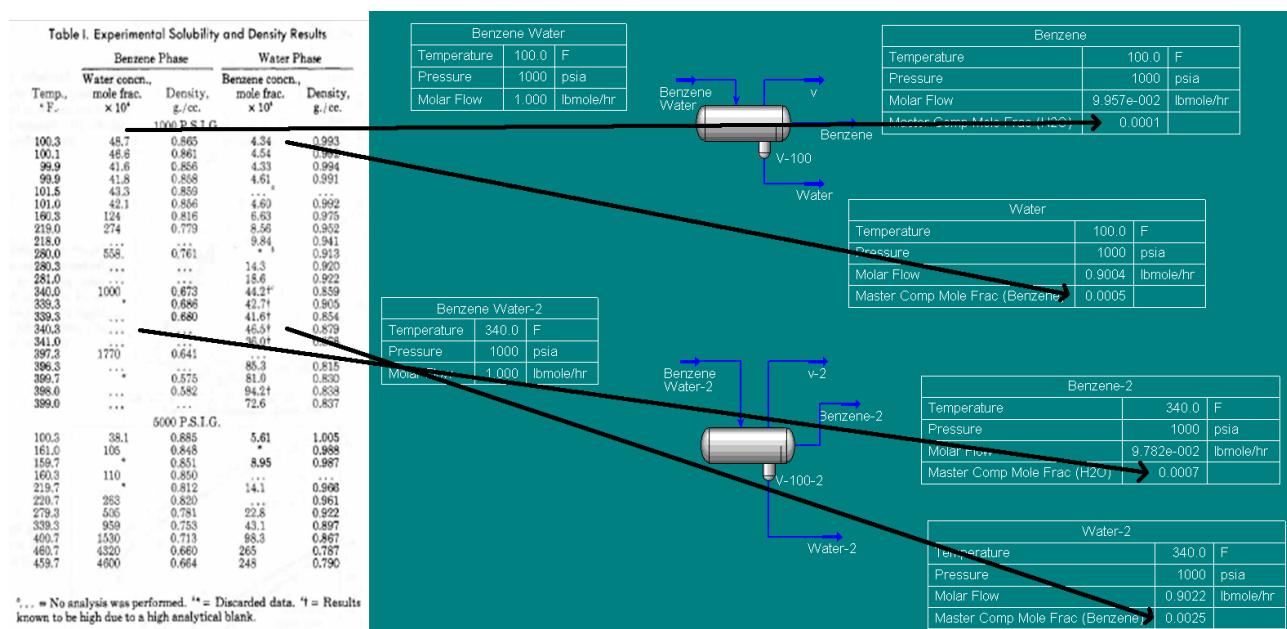
Table. 1 in the attached paper provides data which can be compared with phase behaviour models. A comparison is made on UniSim using Peng Robinson and NRTL (non random two liquid). Benzene and water are combined in a 3 phase separator where the water phase will saturate with benzene and vice versa.

Two data points are compared; 1000 psig (69 bara) at 100 °F and 340 °F (37.8 °C and 171 °C). Peng-Robinson does a very poor job when simulating benzene in water while NRTL, while not exact, is in the correct order of magnitude. We can conclude that Peng-Robinson is unsuitable for this application, an activity model is better suited.

Peng-Robinson



NRTL

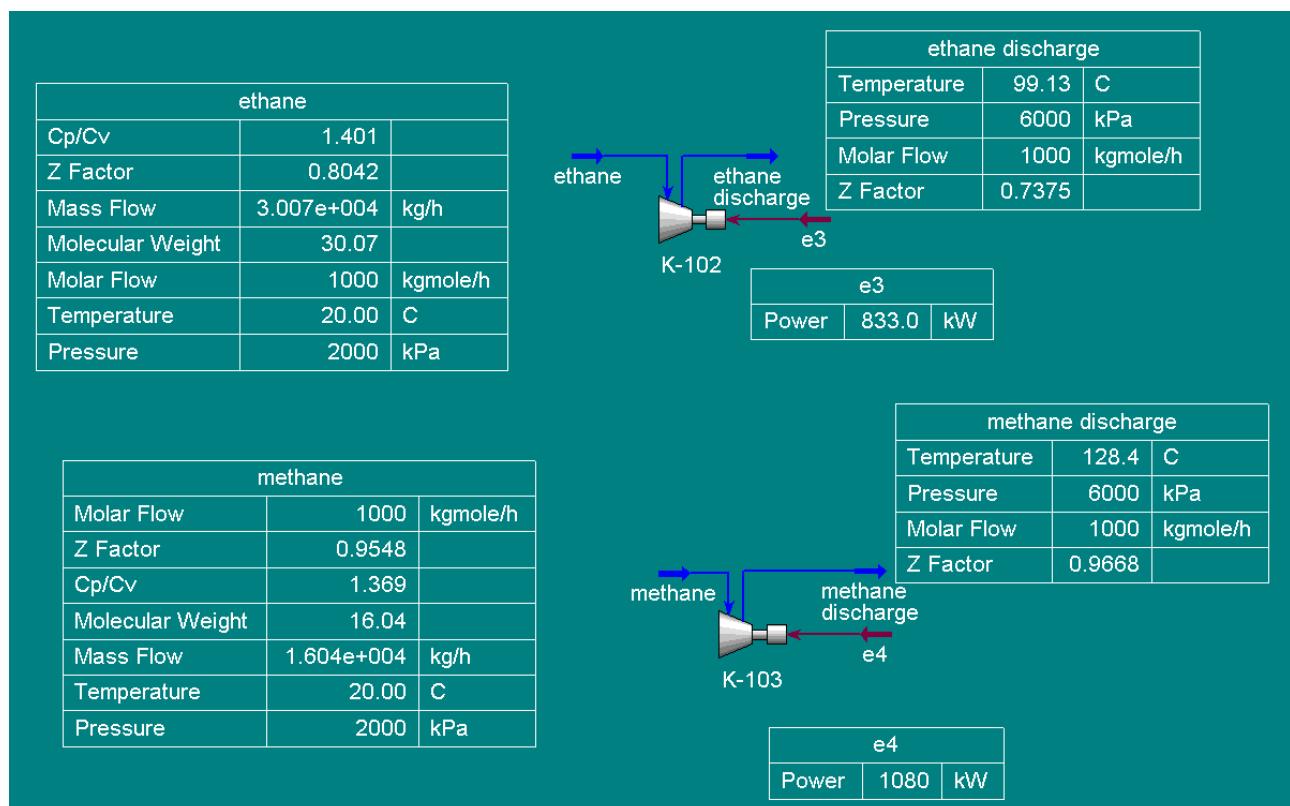


[Question end]

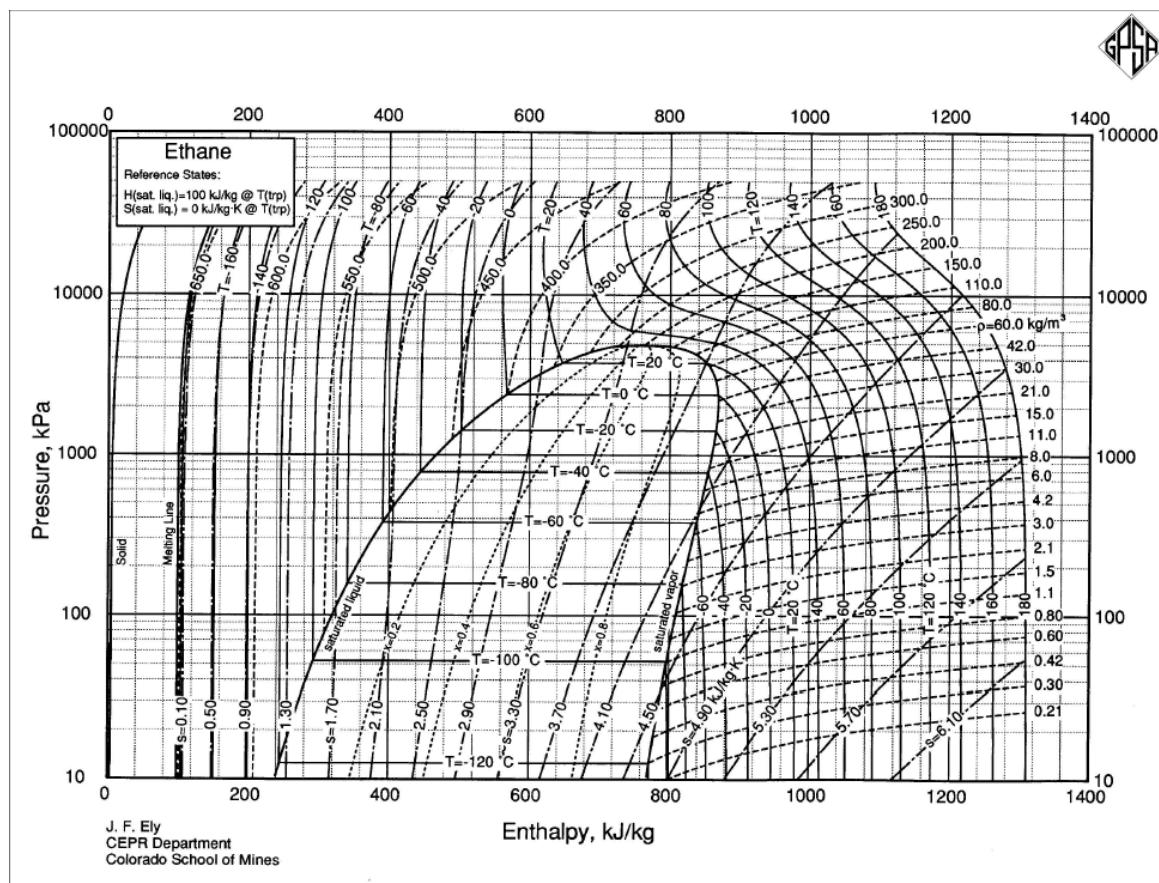
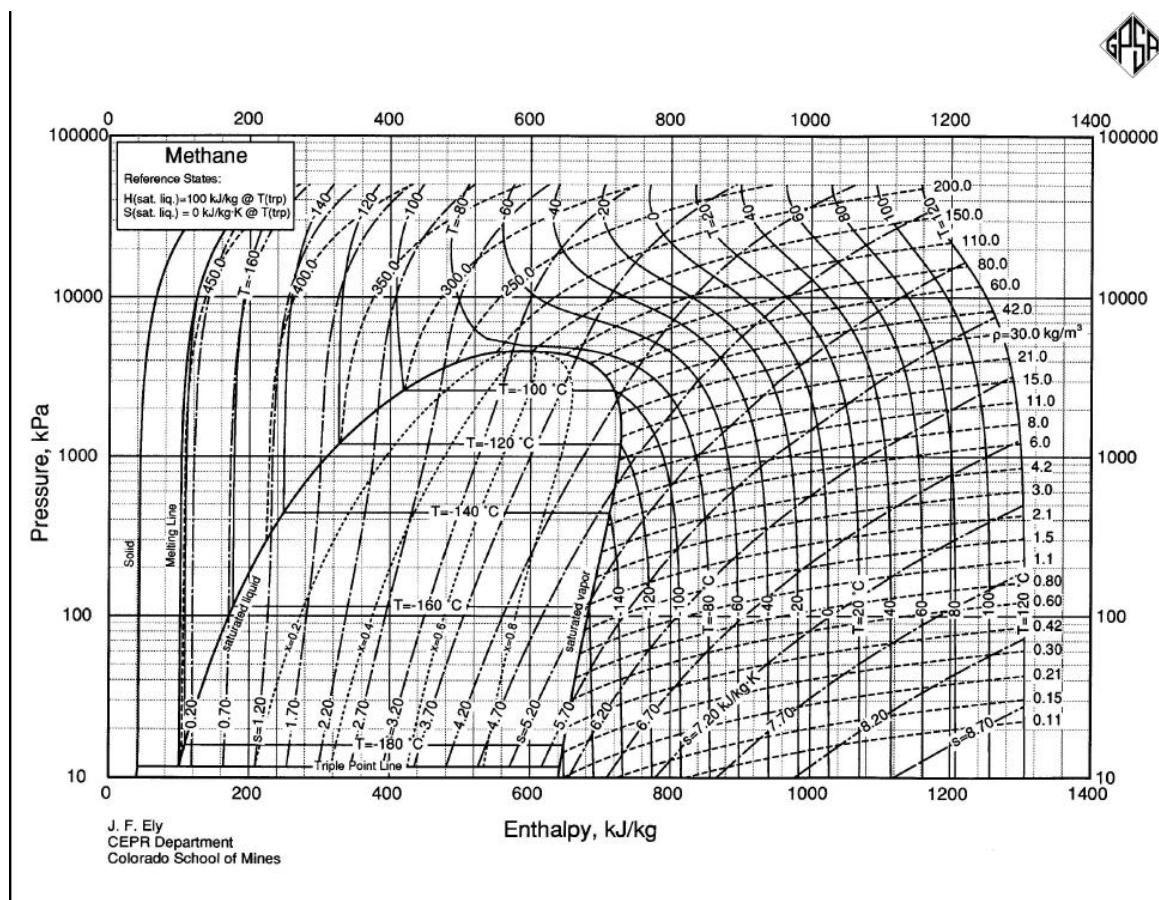
5 5: Gas compression

Q.27 Question 27

The following simulation has been run on UNISIM. Represented are two compressors, one handling 100 % ethane and the other 100% methane. The polytropic efficiency is 77.5% for both compressors.



- Build the same model in UNISIM, ASPEN HYSYS etc.
- Compare the UNISIM predicted power and the compressor outlet temperature with that using the polytropic head approach given in the lecture notes. Do not use the UNISIM C_p/C_V - source the values from literature.
- If the methane compressor outlet temperature rises to 160 °C, what efficiency change would have caused this?
- Predict the power requirements using the attached Mollier charts. Show the compressor path.



Solution:

- Compare the UNISIM predicted power and the compressor outlet temperature with that using the polytropic head approach given in the lecture notes. Do not use the UNISIM C_p/C_v - source the values from literature.

The polytropic head approach given in the lecture notes is shown in the attached image.

Compression Power Sensitivity

- Centrifugal Compressors
 - cover a wide operating range
 - extensive use in upstream oil & gas industry
 - main suppliers
 - Dresser
 - Nuovo Pignone
 - Sulzer
 - good reliability compared to reciprocating machines
 - not suitable for high head low flow applications
 - compression ratio limited by discharge temperature, in turn defined by materials
 - Head equation:

$$\Delta h_{poly} \approx \frac{T_1 \cdot Z_a \cdot R}{\left(\frac{n-1}{n}\right) \cdot MW} \cdot \left[\left(\frac{P_2}{P_1} \right)^{\frac{n-1}{n}} - 1 \right]$$

$$T_2 = T_1 \cdot \left(\frac{P_2}{P_1} \right)^{\frac{n-1}{n}}$$

$$\frac{n-1}{n} \approx \frac{k-1}{k \cdot E_{poly}}$$

For paraffin gases k may be estimated from:

$$k = 1.3 - 0.31 (\gamma - 0.55)$$

| | |
|-------------------|---|
| Δh_{poly} | Polytropic head (kJ/kg) |
| T | Temperature (K) |
| Z_a | Average compressibility ($Z_1 + Z_2$)/2 |
| R | Gas constant (8.314 kJ/kmol.K) |
| P | Pressure (Pa) |
| n | Polytropic coefficient |
| k | Ratio of specific heats (C_p/C_v) |
| MW | Molweight (kg/kmol) |
| W | Gas power (kW) |
| ϕ_m | Mass flow (kg/s) |
| E_{poly} | Polytropic efficiency - Function of volumetric inlet flow varying from approx. 0.6 to 0.8 (compressor specific) |
| γ | Gas relative density (-) |
| $1/2$ | Suction/Discharge |

| Gas | Molecular Weight | Specific Heat Ratio ($k = C_p/C_v$) at 60°F and One Atmosphere | Critical Flow Pressure Ratio at 60°F and One Atmosphere | Specific Gravity at 60°F and One Atmosphere | Critical Constants | |
|------------------|------------------|---|---|--|--------------------|------------------|
| | | | | | Pressure (psia) | Temperature (°F) |
| Methane | 16.04 | 1.31 | 0.54 | 0.554 | 673 | -116 |
| Ethane | 30.07 | 1.19 | 0.57 | 1.058 | 718 | 90 |
| Ethylene | 28.03 | 1.24 | 0.57 ^a | 0.969 | 742 | 50 |
| Propane | 44.09 | 1.13 | 0.58 | 1.522 | 617 | 206 |
| Propylene | 47.08 | 1.15 | 0.58 ^a | 1.453 | 667 | 197 |
| Isobutane | 58.12 | 1.18 | 0.59 ^a | 2.007 | 529 | 273 |
| n-Butane | 58.12 | 1.19 | 0.59 | 2.007 | 551 | 304 |
| 1-Butene | 56.10 | 1.11 | 0.59 ^a | 1.937 | 543 | 276 |
| Isopentane | 72.15 | 1.08 | 0.59 ^a | 2.491 | 483 | 369 |
| n-Pentane | 72.15 | 1.08 | 0.59 ^a | 2.491 | 490 | 386 |
| 1-Pentene | 70.13 | 1.08 | 0.59 ^a | 2.421 | 586 | 377 |
| n-Hexane | 86.18 | 1.06 | 0.59 ^a | 2.973 | 437 | 454 |
| Benzene | 78.11 | 1.12 | 0.58 | 2.697 | 714 | 552 |
| n-Heptane | 100.20 | 1.05 | 0.60 ^a | 3.459 | 397 | 513 |
| Toluene | 92.13 | 1.09 | 0.59 | 3.181 | 590 | 604 |
| n-Octane | 114.22 | 1.05 | 0.60 ^a | 3.944 | 362 | 564 |
| n-Nonane | 128.23 | 1.04 | 0.60 ^a | 4.428 | 552 | 610 |
| n-Decane | 142.28 | 1.03 | 0.60 ^a | 4.912 | 304 | 632 |
| Air | 29.96 | 1.40 | 0.53 | 1.000 | 547 | -221 |
| Ammonia | 17.03 | 1.30 | 0.53 | 0.588 | 1636 | 270 |
| Carbon Dioxide | 44.01 | 1.29 | 0.55 | 1.519 | 1071 | 88 |
| Hydrogen | 2.02 | 1.41 | 0.52 | 0.0696 | 188 | -400 |
| Hydrogen sulfide | 34.08 | 1.32 | 0.53 | 1.176 | 1306 | 213 |
| Sulfur dioxide | 64.04 | 1.27 | 0.55 | 2.212 | 1143 | 316 |
| Steam | 18.01 | 1.33 | 0.54 | 0.622 | 3206 | 706 |

Methane

The polytropic head is calculated

$$\Delta h = \frac{T_1 Z_a R}{\left(\frac{n-1}{n}\right) MW} \left(\left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right) \quad (131)$$

$$= \frac{293 \text{ K} \cdot 0.9608 \cdot 8.314 \frac{\text{J}}{\text{mol K}}}{0.305 \cdot 16 \frac{\text{g}}{\text{mol}}} \left(\left(\frac{60 \text{ bar}}{20 \text{ bar}} \right)^{0.305} - 1 \right) \quad (132)$$

$$= 190.95 \text{ kJ kg}^{-1} \quad (133)$$

and is then used in combination with the mass flow rate in order to calculate the required power

$$P = \dot{m} \Delta h / E \quad (134)$$

$$= \frac{16040 \frac{\text{kg}}{\text{hr}}}{3600 \frac{\text{s}}{\text{hr}}} \cdot 190.95 \frac{\text{kJ}}{\text{kg}} / 0.775 \quad (135)$$

$$= 1097.773 \text{ kW} \quad (136)$$

where \dot{m} is the mass flow rate, and E is the efficiency.

The compressor outlet temperature is given by

$$T_2 = T_1 \left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} \quad (137)$$

$$= 293 \left(\frac{60}{20} \right)^{0.305} \quad (138)$$

$$= 409.78 \text{ K} \quad (139)$$

Ethane

The polytropic head is calculated

$$\Delta h = \frac{T_1 Z_a R}{\left(\frac{n-1}{n}\right) MW} \left(\left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right) \quad (140)$$

$$= \frac{293 \text{ K} \cdot 0.77085 \cdot 8.314 \frac{\text{J}}{\text{mol K}}}{0.206 \cdot 30 \frac{\text{g}}{\text{mol}}} \left(\left(\frac{60 \text{ bar}}{20 \text{ bar}} \right)^{0.206} - 1 \right) \quad (141)$$

$$= 77.17 \text{ kJ kg}^{-1} \quad (142)$$

and is then used in combination with the mass flow rate in order to calculate the required power

$$P = \dot{m} \Delta h / E \quad (143)$$

$$= \frac{30070 \frac{\text{kg}}{\text{hr}}}{3600 \frac{\text{s}}{\text{hr}}} \cdot 77.17 \frac{\text{kJ}}{\text{kg}} / 0.775 \quad (144)$$

$$= 831.715 \text{ kW} \quad (145)$$

where \dot{m} is the mass flow rate, and E is the efficiency.

The compressor outlet temperature is given by

$$T_2 = T_1 \left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} \quad (146)$$

$$= 293 \left(\frac{60}{20} \right)^{0.206} \quad (147)$$

$$= 367.42 \text{ K} \quad (148)$$

- If the methane compressor outlet temperature rises to 160 °C, what efficiency change would have caused this?*

The equation for the compressor output temperature is given by

$$T_2 = T_1 \left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} \quad (149)$$

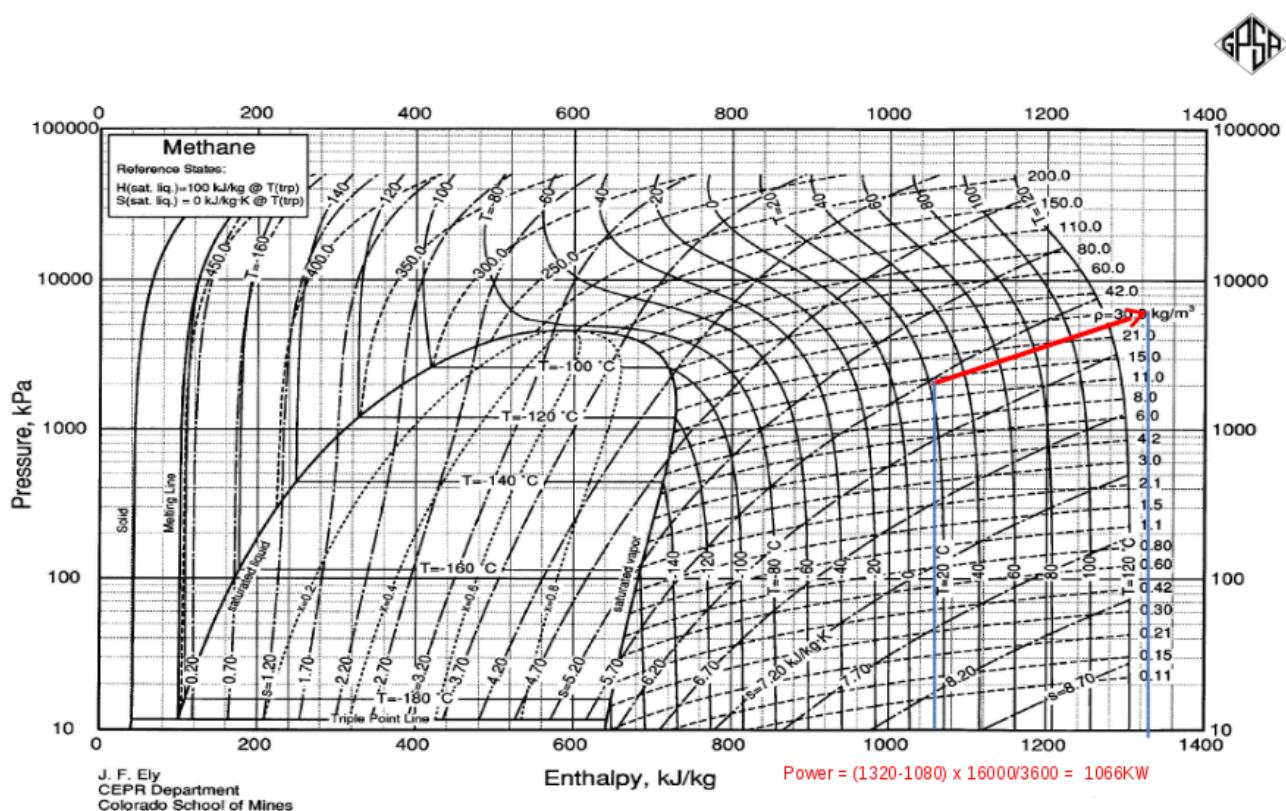
This equation is rearranged for the compressor efficiency E and then used to solve for the E which gives an output temperature of $T = 160 \text{ }^{\circ}\text{C} = 433 \text{ K}$ given that $(n-1)/n \approx (k-1)/kE$

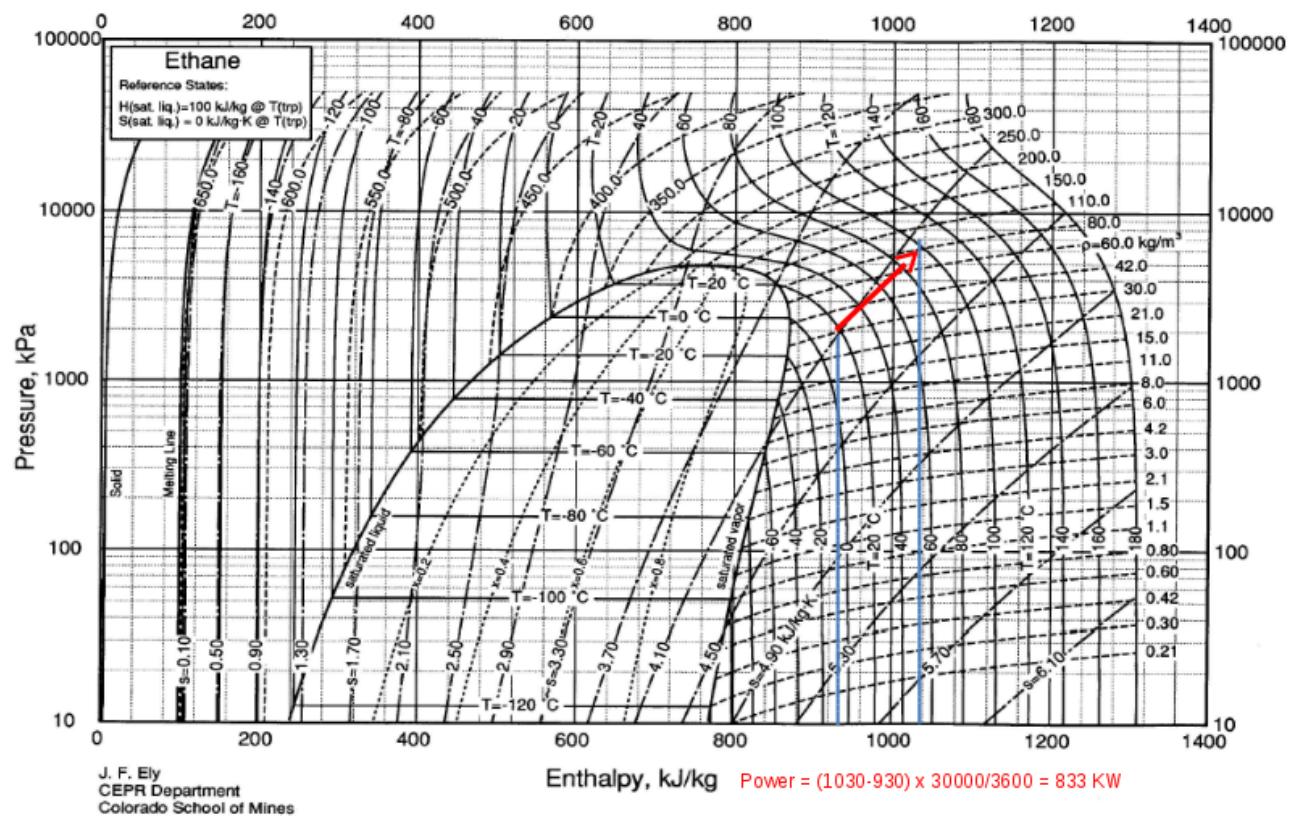
$$E = \frac{k-1}{k} \log_{\frac{T_2}{T_1}} \left(\frac{p_2}{p_1} \right) \quad (150)$$

$$= \frac{1.31-1}{1.31} \log_{\frac{433 \text{ K}}{293 \text{ K}}} \left(\frac{60 \text{ bar}}{20 \text{ bar}} \right) \quad (151)$$

$$= 0.666 \quad (152)$$

- Predict the power requirements using the attached Mollier charts. Show the compressor path.

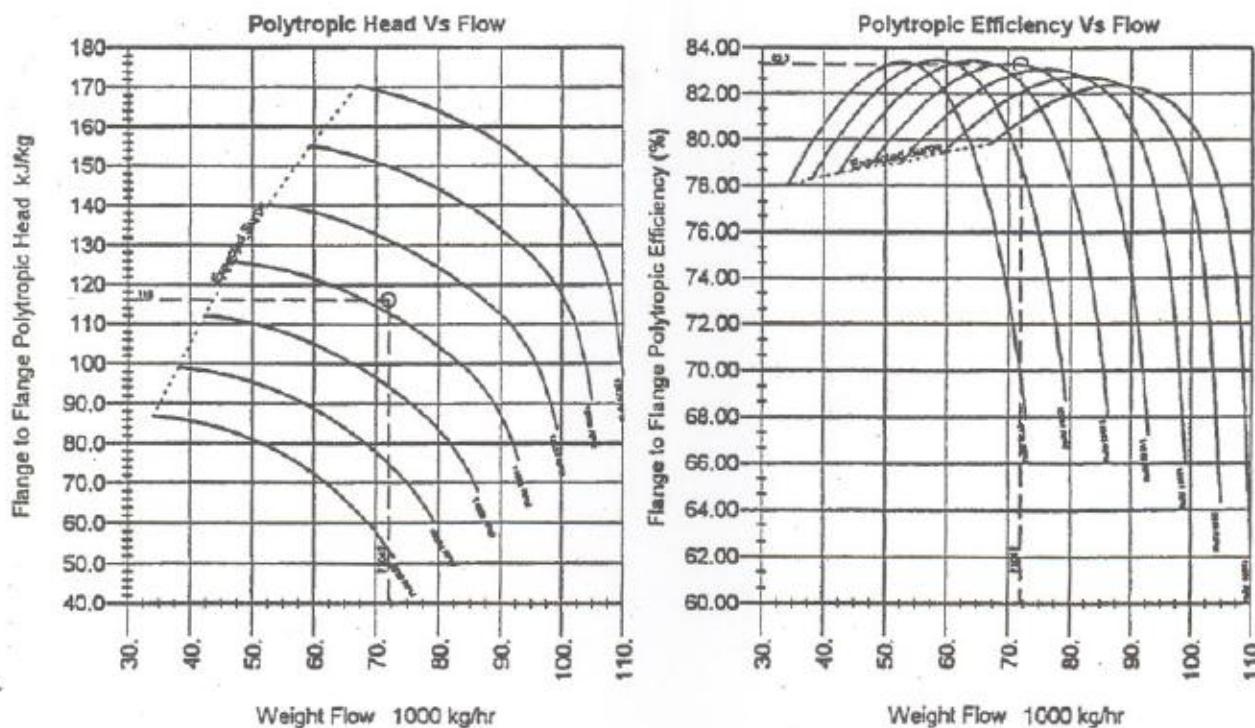




[Question end]

Q.28 Question 28

A compressor has the following performance curves.



- Using the polytropic head expression, what will be the discharge pressure for the following conditions:
 - Suction pressure = 10 bara
 - Suction temperature = 30 °C
 - Molecular weight = 22.6
 - Z = 0.97
 - $C_p/C_v = 1.25$

Use the design point shown on the curves.

- How could you reduce the power requirement (enthalpy change) for the same fluid with the same compression ratio?
- Using the variable you have identified to reduce the enthalpy change, what new value would you change it to in order to reduce the enthalpy change by 5 %?

Solution:

- Using the polytropic head expression, what will be the discharge pressure for the following conditions:
 - Suction pressure = 10 bara
 - Suction temperature = 30 °C
 - Molecular weight = 22.6
 - Z = 0.97

- $C_p/C_V = 1.25$

Use the design point shown on the curves.

(from charts)

- Polytropic head $\Delta h = 116 \text{ kJ kg}^{-1}$
- Polytropic efficiency $E = 83.3 \%$

The polytropic head equation is given as

$$\Delta h = \frac{T_1 Z_a R}{\left(\frac{n-1}{n}\right) MW} \left(\left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right) \quad (153)$$

where $(n - 1)/n \approx (k - 1)/kE$ and $T_1 = 30 \text{ }^\circ\text{C} = 303 \text{ K}$. This equation is rearranged for the discharge pressure and solved

$$p_2 = p_1 \left(\frac{\left(\frac{k-1}{kE}\right) MW \Delta h}{T_1 Z_a R} + 1 \right)^{\frac{kE}{k-1}} \quad (154)$$

$$= 10 \text{ bar} \left(\frac{\left(\frac{1.25-1}{1.25-0.833}\right) 22.6 \frac{\text{g}}{\text{mol}} 116 \frac{\text{J}}{\text{g}}}{303 \text{ K} \cdot 0.97 \cdot 8.314 \frac{\text{J}}{\text{K} \cdot \text{mol}}} + 1 \right)^{\frac{1.25-0.833}{1.25-1}} \quad (155)$$

$$= 26 \text{ bar} \quad (156)$$

- *How could you reduce the power requirement (enthalpy change) for the same fluid with the same compression ratio?*

Consider the polytropic head equation

$$\Delta h = \frac{T_1 Z_a R}{\left(\frac{n-1}{n}\right) MW} \left(\left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right). \quad (157)$$

Reducing the temperature will reduce the head and therefore the power.

- *Using the variable you have identified to reduce the enthalpy change, what new value would you change it to in order to reduce the enthalpy change by 5 %?*

Temperature is directly proportional to power $T \propto \Delta h \propto P$ in this case so a 5% drop in temperature will result in a 5% drop in power.

$$T_{new} = (1 - 0.05) T_{old} \quad (158)$$

$$= 0.95 \cdot 303 \text{ K} \quad (159)$$

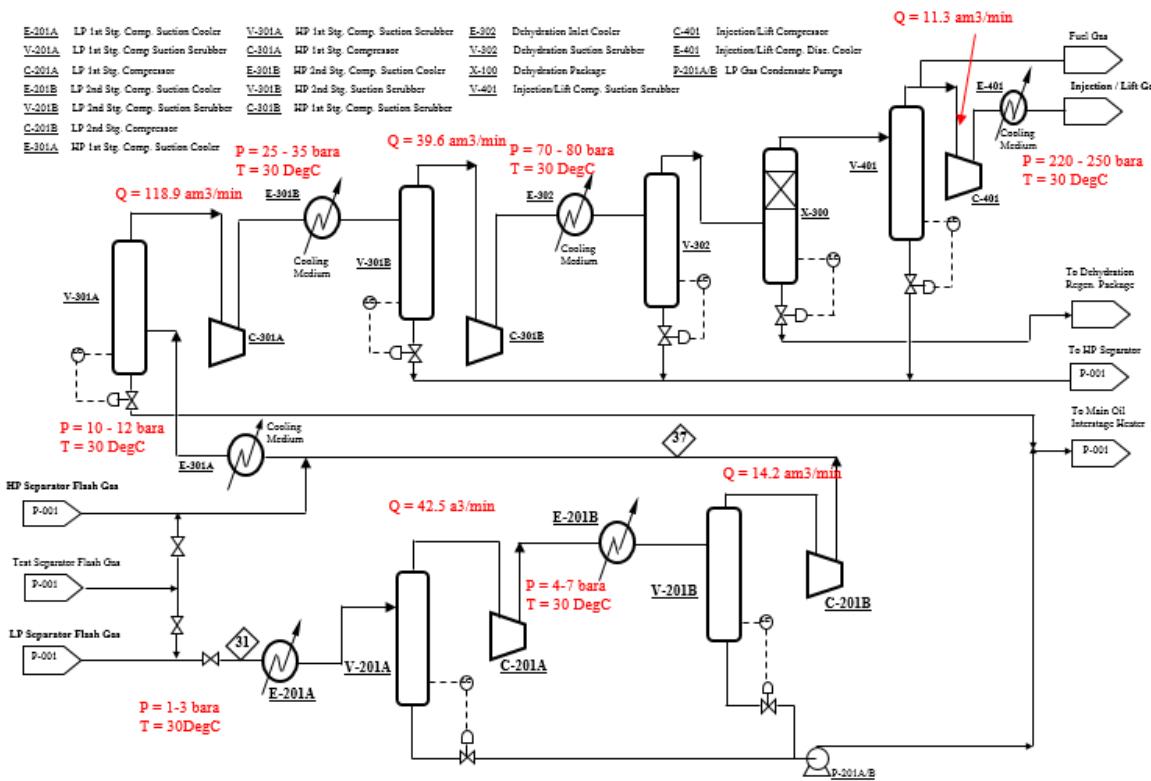
$$= 288 \text{ K} = 15 \text{ }^\circ\text{C} \quad (160)$$

[Question end]

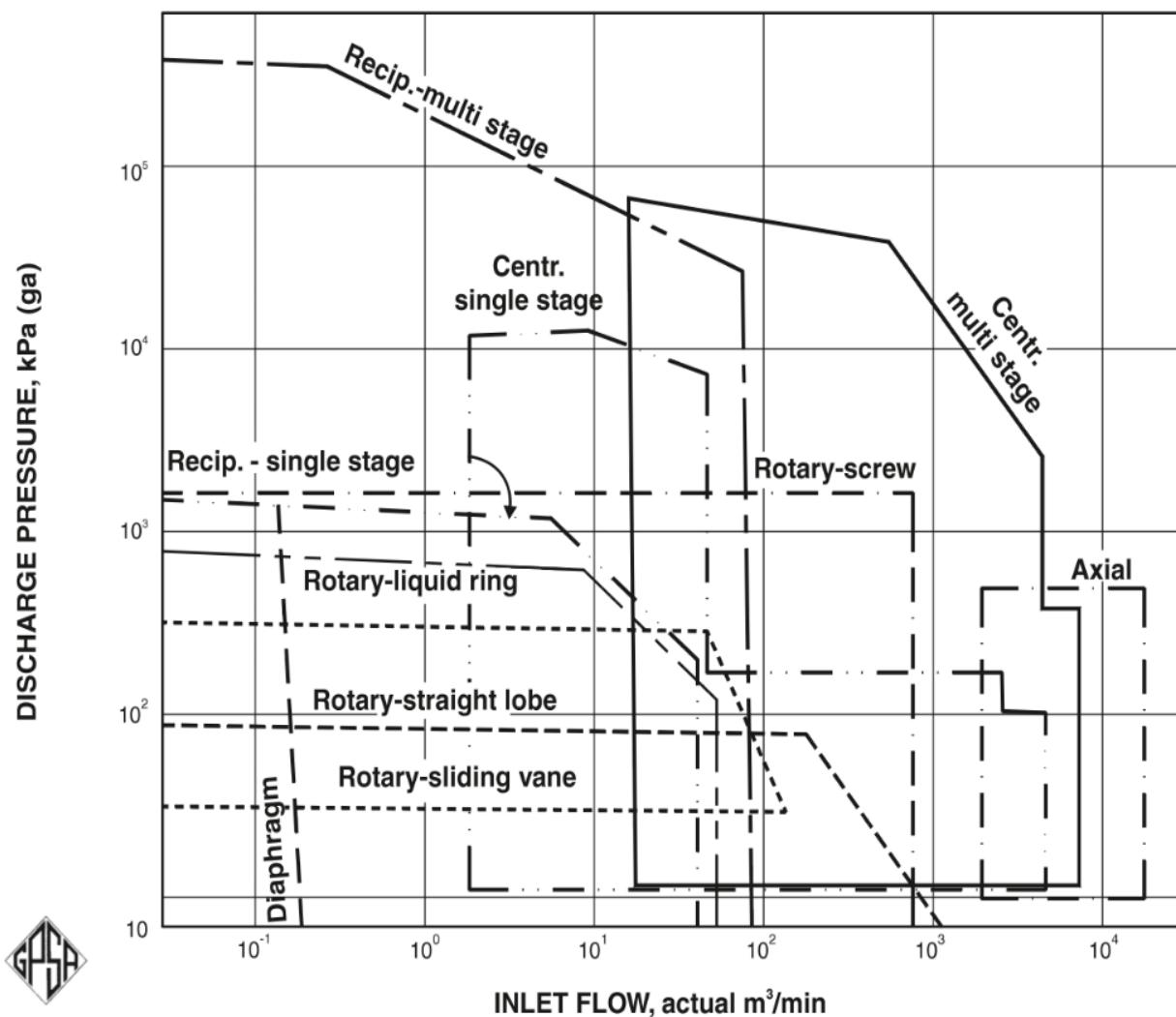
Q.29

Question 29

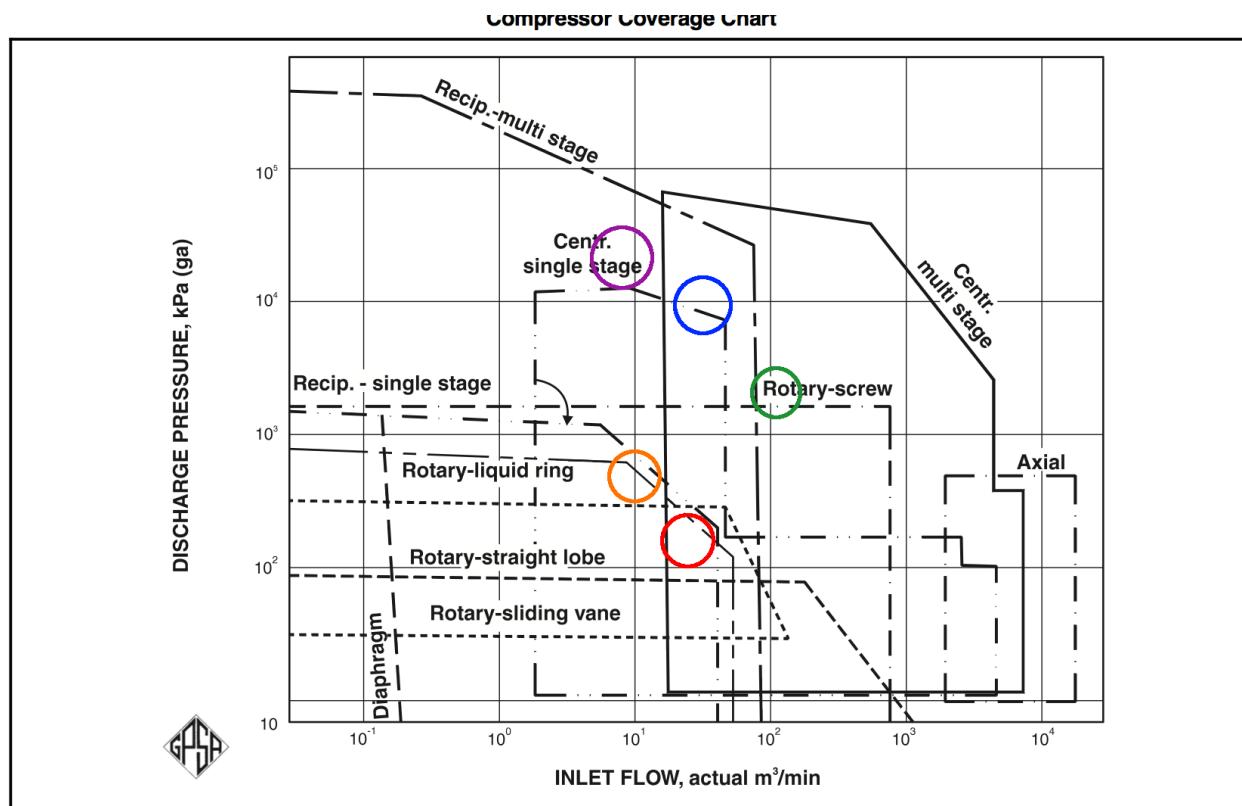
The attached PFD shows a compression train.

Gas Plant PFD

Overlaid on to it are the compressor inlet flow rates as well as the compressor outlet conditions. Use this information and the compressor selection map to identify compressor types for the various duties.



Solution:



- ■ C201 A - Screw
- ■ C201 B - Screw
- ■ C301 A - Cent.
- ■ C301 B - Cent.
- ■ C301 C - Recip.

[Question end]

Q.30 Question 30

5: Gas compression

- a) Name and briefly describe four types of compressors including a description of their operating principles. [8 marks]

Solution:

Any four of the below are accepted:

Centrifugal compressors use a rotating disk or impeller in a shaped housing. Gas enters the impeller and is moves to the rim of the impeller, increasing the velocity of the gas. A diffuser (divergent duct) section converts the velocity energy to pressure energy. [2/8]

Axial-flow compressors are dynamic rotating compressors that use arrays of fan-like airfoils to progressively compress the working fluid. The arrays of airfoils are set in rows, usually as pairs: one rotating and one stationary. The rotating airfoils, also known as blades or rotors, accelerate the fluid. The stationary airfoils, also known as stators or vanes,

[2/8] decelerate and redirect the flow direction of the fluid, preparing it for the rotor blades of the next stage.[✓]

[2/8] **Reciprocating compressors** use pistons driven by a crankshaft. The piston sits within a cylinder with inlet and exhaust valves. On the down stroke the inlet valve opens allowing gas to enter the cylinder. On the upstroke the gas pressure is increased with the gas being realised at a set pressure through the exhaust valve.[✓]₂

[2/8] **Rotary process screw compressors** operate by drawing gas into the spaces between the lobes of the twin screws. As the rotors turn, the gas is forced by the profile of the screws into a continuously decreasing space until it reaches the outlet port at high pressure.[✓]₂

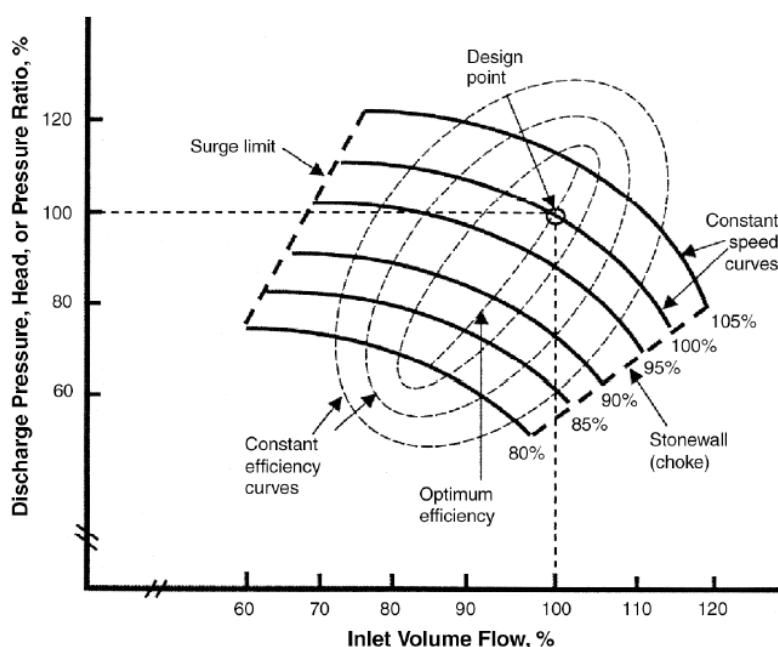
Jet pumps use a high pressure stream to generate a fast stream or jet that is injected into another stream to transfer kinetic energy which is then converted to pressure again as the two mixed streams decelerate.

Liquid ring seal compressors This is similar to a axial flow or reciprocating compressor in that it isolates a volume of gas then constricts the volume to increase the pressure before discharge, but it uses a liquid ring to provide the sealing between compressed volumes (and to provide the constriction).

- b) Sketch a typical head flow characteristic for a centrifugal compressor. Indicate the effect of varying the compressor speed and the surge limit. [4 marks]

Solution:

Marks are awarded based on the detail of the figure[✓]₄



- c) A centrifugal gas compressor is operating with the following inlet and outlet conditions:
- Inlet pressure, $p_1 = 10$ bara.

- Outlet pressure, $p_2 = 30$ bara.
- Inlet temperature, $T_1 = 30^\circ\text{C}$.
- Outlet temperature, $T_2 = 110^\circ\text{C}$.

After a process upset the compressor achieves the same outlet pressure but the outlet temperature of the compressor is seen to rise by 5°C . It is suspected that the machine is losing polytropic efficiency. Calculate the percentage change in the compressor efficiency from before to after the upset.

The compression relationship is as follows;

$$T_2 = T_1 \left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} \quad \frac{n-1}{n} = \frac{k-1}{k \eta_{poly}}$$

where k is the ratio of specific heats for the gas and η_{poly} is the compressor polytropic efficiency. [8 marks]

Solution:

We use the compressor temperature relationship to calculate $(n - 1)/n$ before the upset:

$$\frac{n-1}{n} = \ln(T_2/T_1) / \ln(P_2/P_1) \quad (161)$$

$$= \ln(383/303) / \ln(3/1) \quad (162)$$

$$= 0.213 \quad (163)$$

[2/8] ✓ Calculating the value after upset,

$$\frac{n-1}{n} = \ln(T_2/T_1) / \ln(P_2/P_1) \quad (164)$$

$$= \ln(388/303) / \ln(3/1) \quad (165)$$

$$= 0.225 \quad (166)$$

[2/8] ✓ Since the gas has not changed composition, k will remain constant, hence the change in the polytropic efficiency will be

$$0.213/0.225 = 0.947 \quad (167)$$

I.e. a 5.3% reduction in efficiency.

[Question total: 20 marks]

6 6: Gas treatment

Q.31 Question 31

a) **3: Separators** Stokes law is given by,

$$v_{settle} = \frac{d^2(\rho_w - \rho_o)g}{18\mu}$$

For a system with water droplets separating from a continuous oil phase, name the variables in the equation. [2 marks]

Solution:

- v_{settle} = droplet settling velocity.
- d = droplet diameter
- ρ_w = water density.
- ρ_o = oil density.
- μ = oil viscosity.
- g = gravitational constant.

[2/2]

✓
2**b) 3: Separators**

An oil, gas, water separator is producing an oil product which meets the required water specification. The oil density is 750 kg m^{-3} and its viscosity is 1 cp while the water density is 1000 kg m^{-3} . As a result of introducing a new oilfield, the oil density and viscosity increase to 850 kg m^{-3} and 1.5 cp respectively. Assuming the droplet size distribution of water in oil stays the same, what percentage change would be required to the oil residence time to achieve the same water specification? [5 marks]

Solution:

Assuming Stokes law holds, the original situation is,

$$\begin{aligned} v_{settle} &= \frac{d^2(\rho_w - \rho_o)g}{18\mu} \\ &= \frac{d^2(1000 - 850)}{18 \cdot 2}. \end{aligned}$$

[1/5]

✓ (Note the units are not exactly right but they cancel so its not important) For the new state,

$$v_{settle} = \frac{d^2(1000 - 900)}{18 \cdot 4}$$

[1/5]

✓ Dividing the two equations the reduction in settling velocity for the same droplet size is obtained,

$$(100/4)/(150/2) = 1/3.$$

[1/5]

✓ Thus a droplet will settle at 1/3rd of the original velocity therefore the residence time must treble to achieve the same degree of separation ✓ .

[2/5]

c) 4: Produced Water

A hydrocyclone is used to remove oil droplets from a continuous water phase. Describe the operating principles of a hydrocyclone. [6 marks]

Solution:

The hydrocyclone consists of four sections: a cylindrical swirl chamber, a concentric reducing section, a fine tapered section, and a cylindrical tail section. The oily water enters the cylindrical swirl chamber through a tangential inlet, creating a high-velocity vortex with a reverse-flowing central core. The fluid accelerates as it flows through the concentric reducing section and the fine tapered section. The fluid then continues at a constant rate through the cylindrical tail section. Larger oil droplets are separated out from the fluid in the fine tapered section, while smaller droplets are removed in the tail section. Centripetal forces cause the lighter-density droplets to move toward the low pressure central core, where axial reverse flow occurs. The oil is removed through a small-diameter reject port located in the head of the hydrocyclone separator. Clean water is removed through the downstream outlet ✓ .

[6/6]

d) 5: Gas Compression

In order to obtain increased discharge pressure from a centrifugal compressor it has been proposed to run the machine at reduced throughput. Why would a reduced throughput result in an increased discharge pressure? What machine characteristic would you check before agreeing to conduct a reduced throughput trial? Why is this characteristic a concern?

[3 marks]

Solution:

The head/pressure flow characteristics of a centrifugal compressor are declining head with increased rate. Hence reducing rate will increase head/pressure₁.

Check for surge. Centrifugal compressors can surge at low flows and high discharge pressures. Surge is a condition of minimum flow/maximum head where the developed head is insufficient to overcome the system resistance. When the compressor reaches this point, the gas in the discharge piping back-flows into the compressor. The repeated pressure oscillations at the surge point should be avoided since it can be detrimental to the compressor₂.

[2/3]

e) 6: Gas Treatment - refrigeration

In order to recover additional hydrocarbon liquids from a gas stream,a refrigeration plant is often utilised to cool the fluid below that possible by air or cooling water. Describe a typical refrigeration cycle. A sketch may help.

[4 marks]

Solution:

A circulating refrigerant such as Freon enters a compressor as a vapour. The vapour is compressed and exits the compressor superheated. The superheated vapour passes to a condenser which removes the superheat and then condenses the vapour at essentially constant pressure and temperature. The pressurised liquid refrigerant is expanded across a throttle valve where Joule-Thomson cooling takes place. The cold liquid-vapour mixture is passed to an evaporator (process cooler) where it is vapourised at a low constant temperature. The evaporator cools the process fluid - the latent heat of vapourisation of the refrigerant is supplied by extracting heat from the process fluid, hence reducing the temperature of the process fluid. The resulting refrigerant vapour returns to the compressor inlet to complete the thermodynamic cycle₄. 1 mark is given for each component of the cycle compressor, condenser, J-T valve, and evaporator.

[4/4]

[Question total: 20 marks]

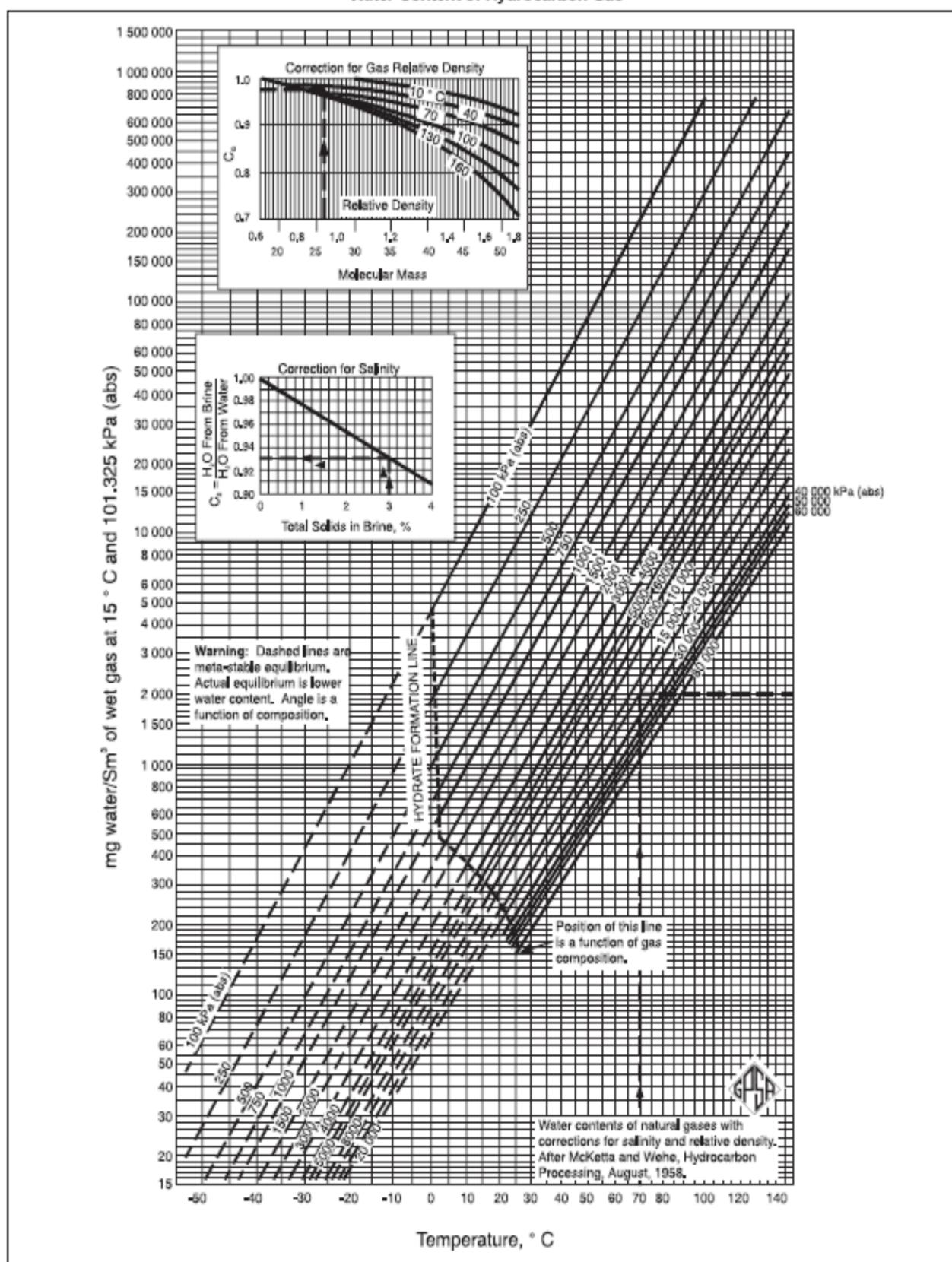
Q.32

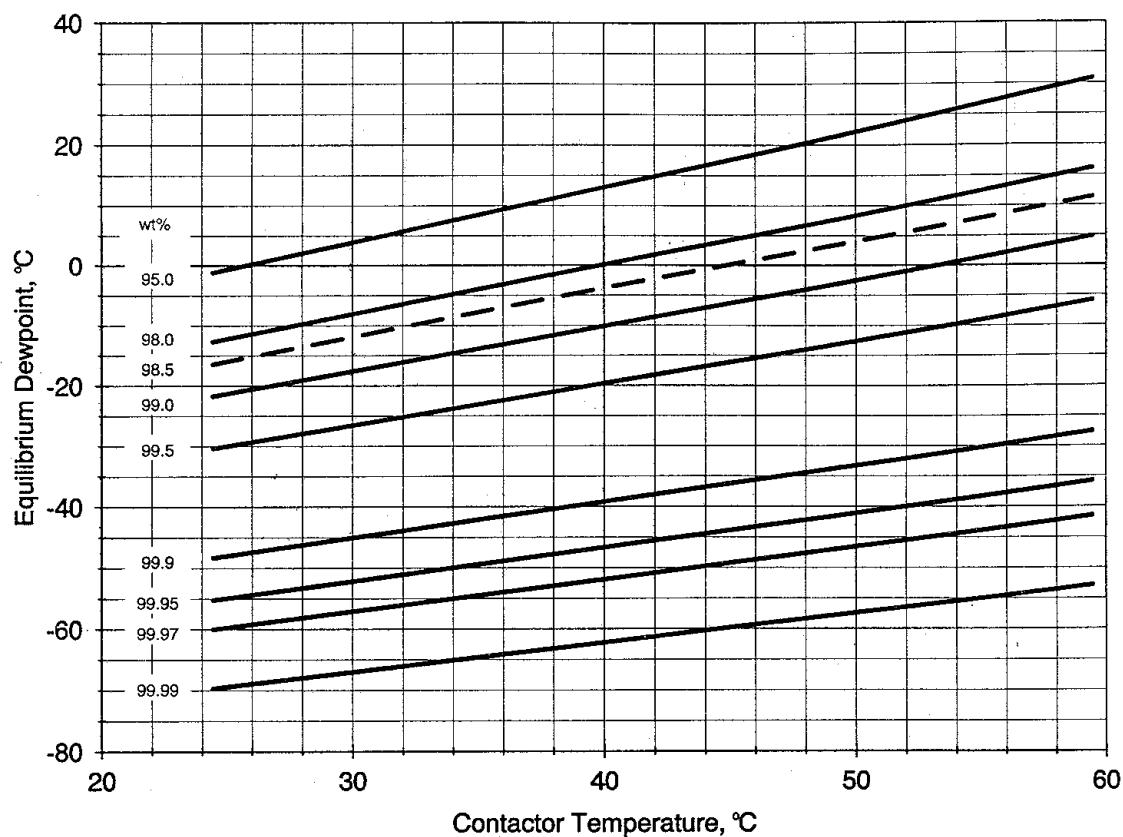
Question 32

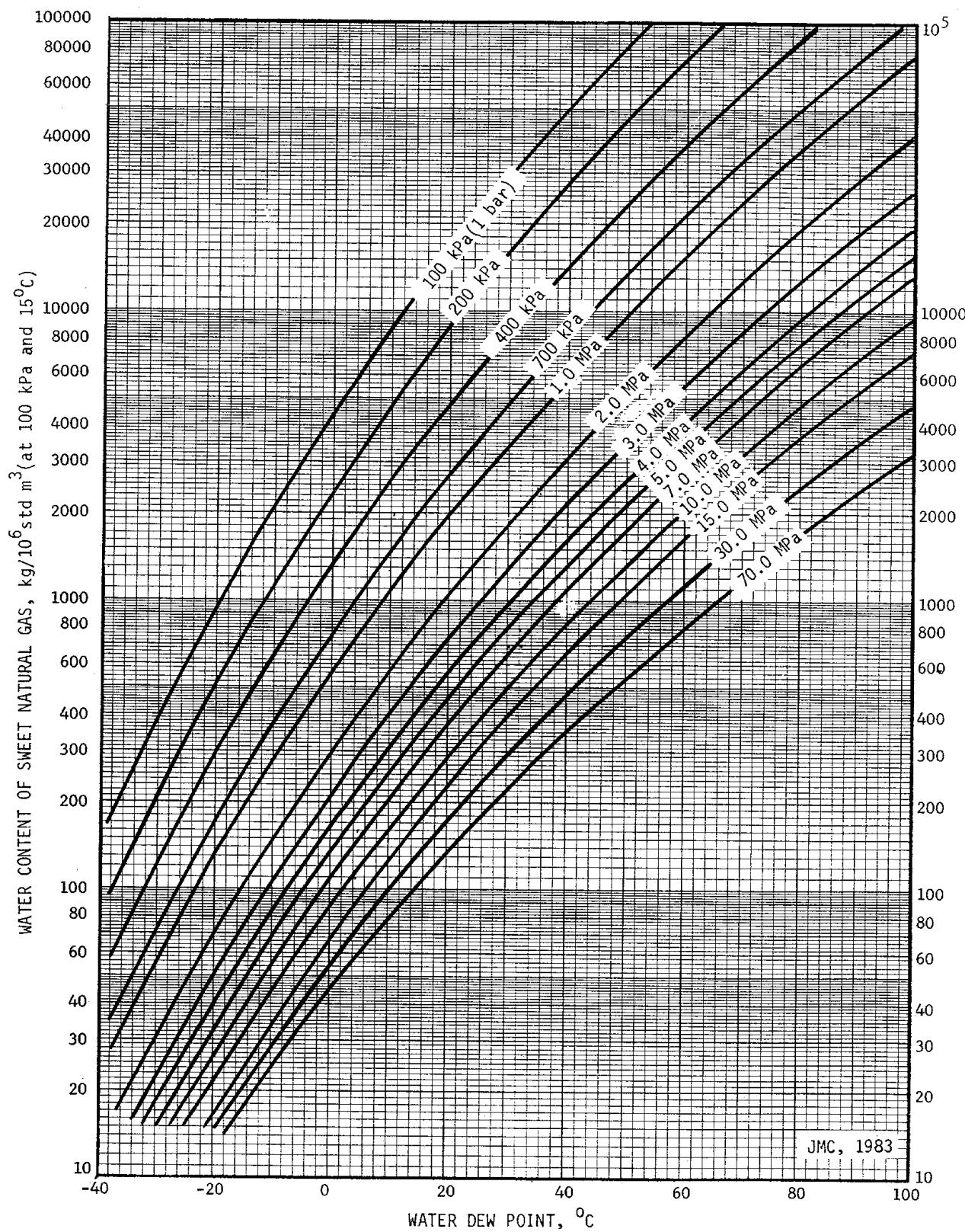
Water saturated methane gas enters a tri-ethylene glycol counter-flow absorber at 30 °C and 50 bara. The duty of the absorber is to dry the gas to a specification of 100 mg H₂O sm⁻³. The flowrate of the gas is 5×10^6 sm³ day⁻¹. Calculate the number of theoretical and actual plates required to achieve the above duty. Use the enclosed water saturation and glycol-water dewpoint equilibrium charts for your calculations. Use the guidance for lean glycol rate and concentration in the lecture notes.

Assume that the column pressure drop is negligible and that the column operates isothermally. Plate efficiencies can be taken as 50 %. Assume a TEG density of 1110 kg m⁻³.

Water Content of Hydrocarbon Gas







Solution:

We know the absorber pressure $p = 50$ bara and temperature $T = 30^\circ\text{C}$ as well as the outlet gas water concentration $c_{H_2O} = 100 \text{ mg H}_2\text{O sm}^{-1}$ and the gas flowrate $Q = 5 \times 10^6 \text{ sm}^3 \text{ day}^{-1}$.

1. The inlet gas is water saturated at 50 bar and 30 °C. Hence we can use the water content chart to determine the inlet gas water content to be 800 mg sm⁻³.
2. The outlet gas specification is 100 mg sm⁻³. Determine the water dewpoint of the outlet gas from the water content chart = -5 °C. To set up a driving force for mass transfer (as equilibrium will never be reached) assume a further 10 °C difference i.e. -15 °C.
3. Calculate the glycol concentration that is required for equilibrium dewpoint of -15 °C and absorber temperature of 30 °C. Using the glycol equilibrium chart - 98.7 wt%.
4. 98.7 wt% glycol will be in equilibrium with a gas with a water dewpoint of -15 °C. From the water content chart this equates to 48 mg sm⁻³.
5. Calculate how much water is transferred to the glycol.

$$Q_{H_2O,transferred} = Q_{gas} \times (c_{H_2O,in} - c_{H_2O,out}) \quad (168)$$

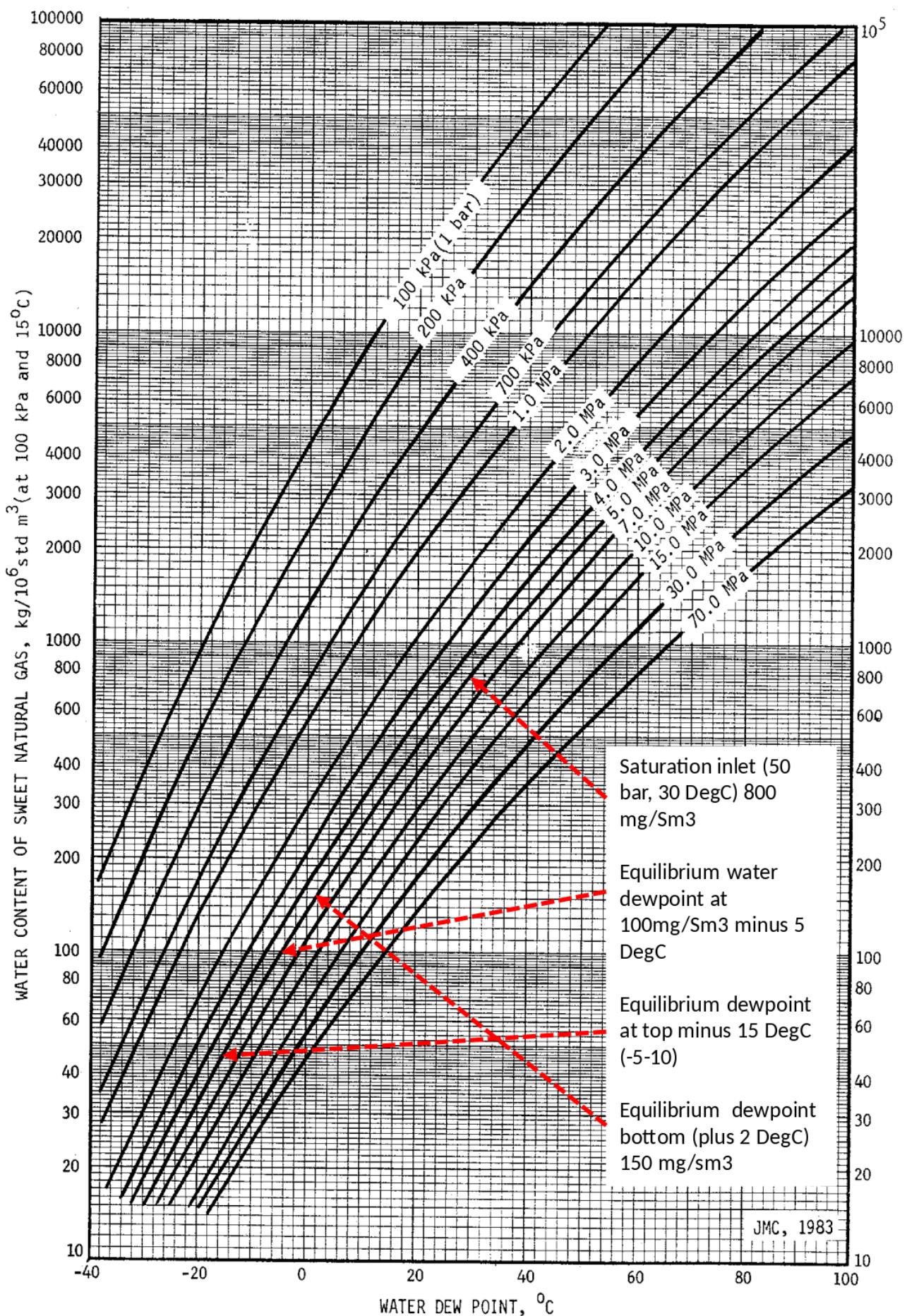
$$= 0.0405 \text{ kg s}^{-1} \quad (169)$$

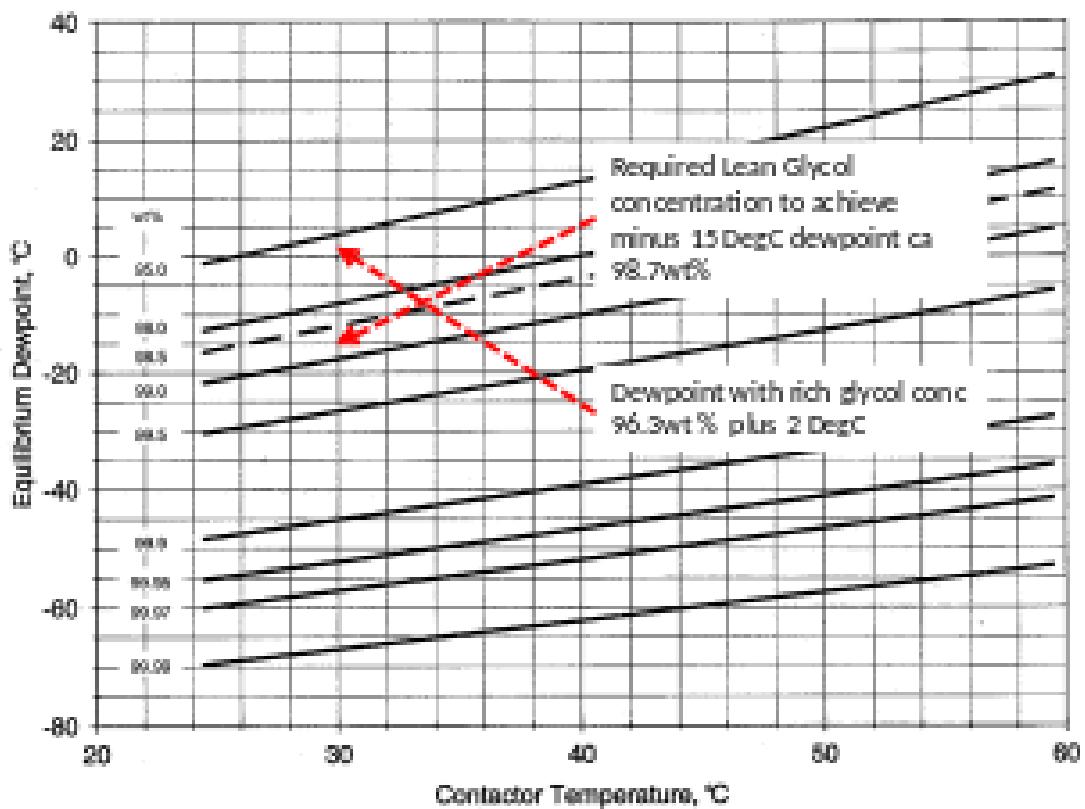
6. Use lecture notes to estimate a glycol rate (the notes on the “Glycol Circulation Rate” indicate that beyond 0.03 m³ (kg H₂O)⁻¹ there is no significant benefit).
7. Calculate total glycol rate,

$$Q_{glycol} = 0.0405 \frac{\text{kg H}_2\text{O}}{\text{s}} \cdot 0.03 \frac{\text{m}^3 \text{ glycol}}{\text{kg H}_2\text{O}} \quad (170)$$

$$= 0.00122 \frac{\text{m}^3 \text{ glycol}}{\text{s}} = 1.35 \frac{\text{kg glycol}}{\text{s}} \quad (171)$$

8. Calculate the outlet concentration of glycol having absorbed 0.0405 kg s⁻¹ of water - 95.8 wt%.
9. Calculate equilibrium water dewpoint at 95.8 wt% and a contactor at 30°C using the glycol equilibrium chart = 2 °C.
10. Calculate the water content in the gas at 2 °C and 50 bar = 150 mg sm⁻³. This is the equilibrium content of water in a gas that is in contact with glycol at 95.8 wt%.
11. Construct the operating and equilibrium lines and step off the theoretical plates and 50% efficient plates (shown in attached drawing).





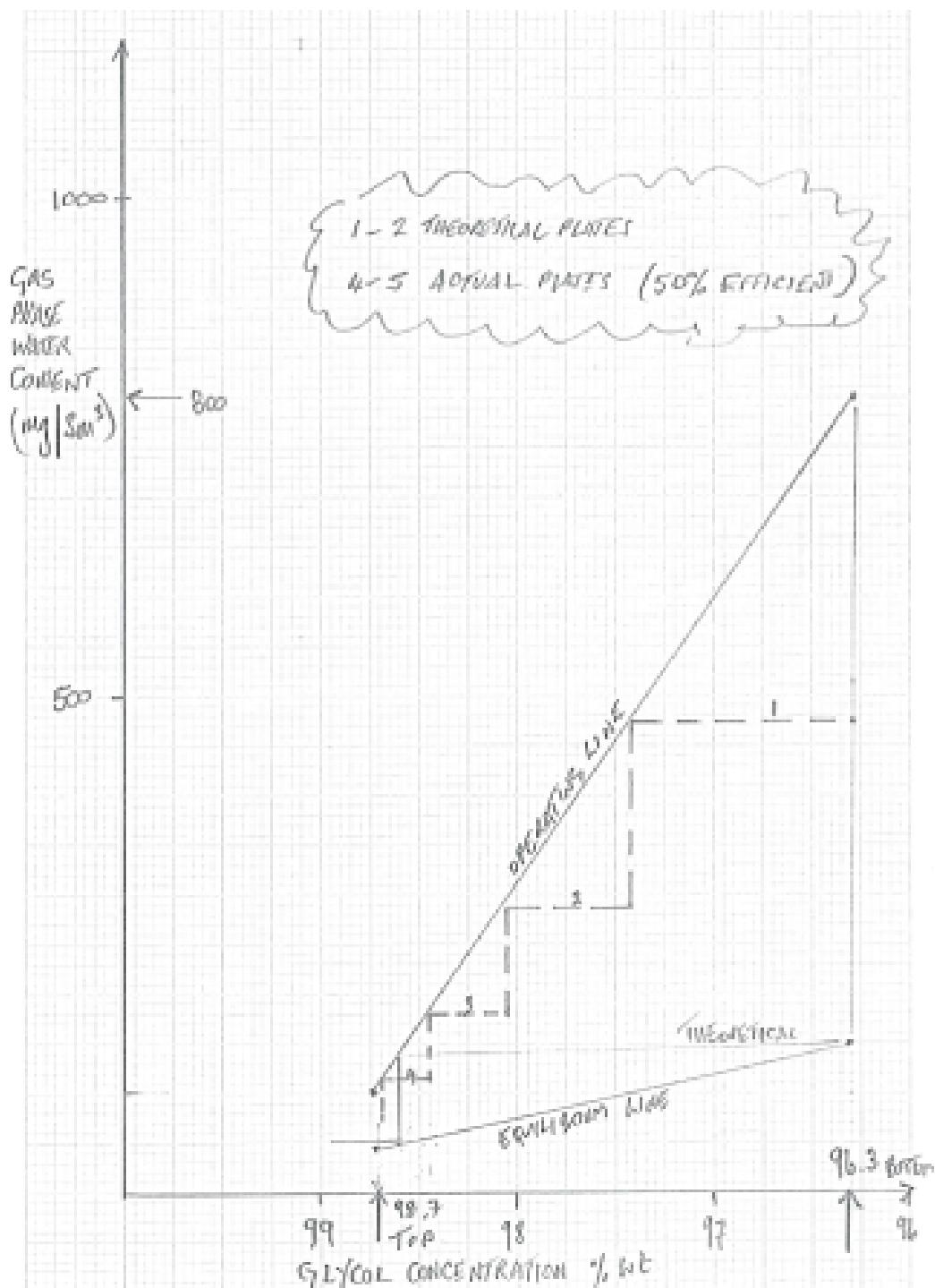
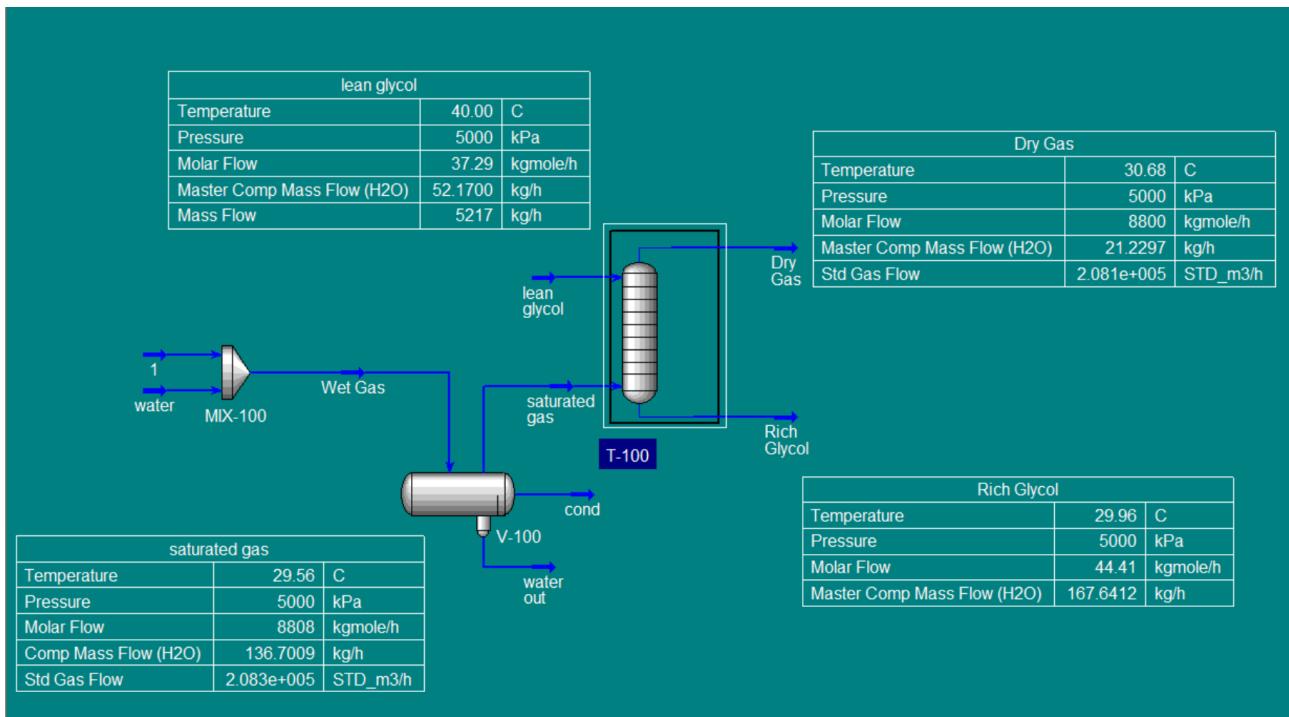


Figure 5: This graph uses an incorrect bottoms concentration of 96.3%, when it should be 95.8% but the values do not change significantly.

[Question end]

Q.33**Question 33**

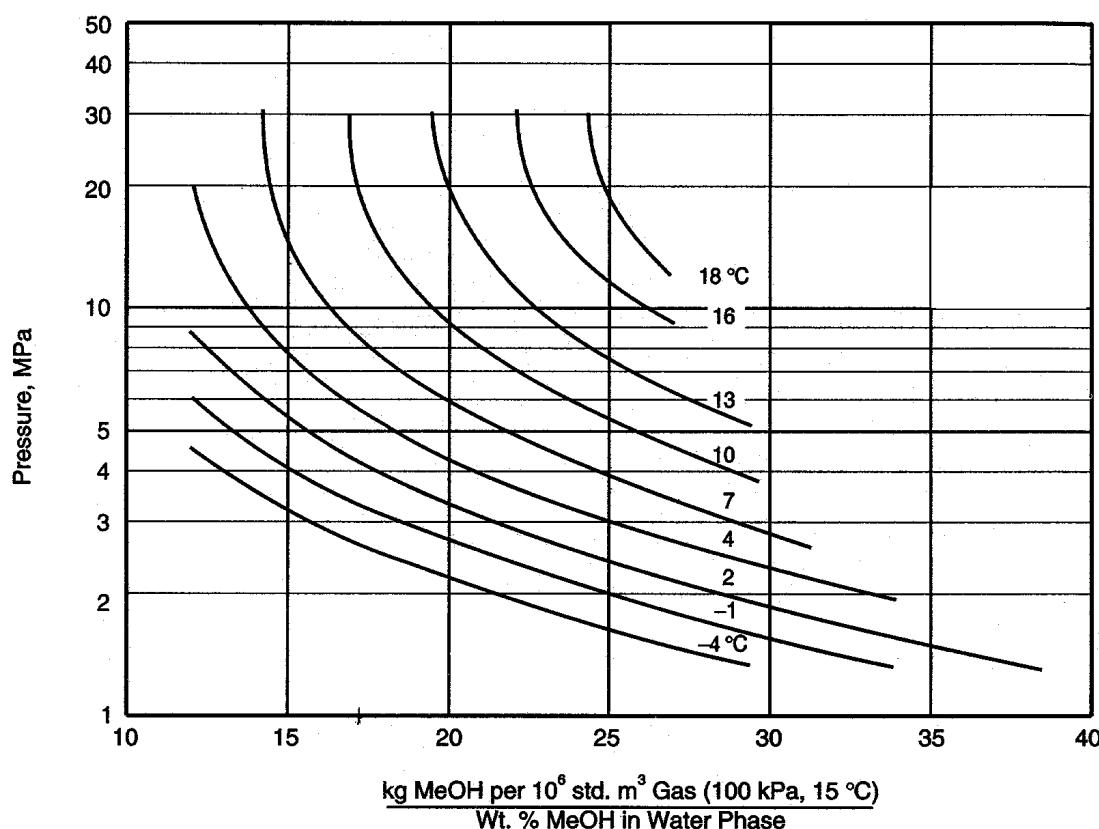
Build a glycol model on UNISIM, ASPEN HYSYS, etc. with the following conditions and compare the answers.



[Question end]

Q.34**Question 34**

10 MMscfd ($2.83 \times 10^5 \text{ stm}^3 \text{ day}^{-1}$) of natural gas ($\gamma = 0.65$) having a hydrate formation temperature of 17°C cools to 4.4°C in a buried pipeline (pressure is 8 MPa). To avoid hydrate formation, how much methanol ($\rho = 800 \text{ kg m}^{-3}$, MW = 32) must be added if the gas enters the line saturated with water at 30°C .



Solution:

Calculate the hydrate point depression

$$\Delta T = 17 \text{ }^{\circ}\text{C} - 4.4 \text{ }^{\circ}\text{C} = 12.6 \text{ }^{\circ}\text{C} \quad (172)$$

Use the water content chart for sweet natural gas to calculate the water content at

- $30 \text{ }^{\circ}\text{C} = 540 \frac{\text{kg}}{10^6 \text{sm}^3}$

- $4.4 \text{ }^{\circ}\text{C} = 140 \frac{\text{kg}}{10^6 \text{sm}^3}$

The total amount of liquid H_2O is therefore

$$Q_{\text{liqH}_2\text{O}} = 400 \frac{\text{kg H}_2\text{O}}{10^6 \text{sm}^3} \cdot 0.283 \times 10^6 \frac{\text{sm}^3}{\text{day}} = 113 \frac{\text{kg H}_2\text{O}}{\text{day}} \quad (173)$$

Use the Hammerschmidt equation to find the wt% of methanol required in water

$$x_{\text{MeOH}} = 100 \cdot \frac{12.6 \cdot 32}{1297 + 12.6 \cdot 32} = 23.7 \text{ wt\%} \quad (174)$$

Calculate the required mass of methanol in water

$$m_{\text{MeOH}} = 113 \frac{\text{kg H}_2\text{O}}{\text{day}} \cdot \frac{23.7 \text{ wt\% H}_2\text{O}}{100 - 23.7 \text{ wt\% H}_2\text{O}} = 35.1 \frac{\text{kg MeOH}}{\text{day}} \quad (175)$$

Using the attached vapour equilibrium chart, determine the methanol fraction lost to the vapour phase

$$\text{MeOH Vap. losses (from chart)} = 15 \frac{\text{kg MeOH}/10^6 \text{sm}^3}{\text{wt\% MeOH in water}} \quad (176)$$

Use this information to calculate the mass rate of methanol lost

$$m_{\text{MeOH}} = 15 \frac{\text{kg MeOH}/10^6 \text{sm}^3}{\text{wt\% MeOH in water}} \cdot 0.283 \times 10^6 \frac{\text{sm}^3}{\text{day}} \cdot 23.7 \text{ wt\% MeOH} \quad (177)$$

$$= 100.6 \frac{\text{kg MeOH}}{\text{day}} \quad (178)$$

The total required injection rate of methanol is therefore

$$m_{\text{MeOH},\text{total}} = 35.1 \frac{\text{kg MeOH}}{\text{day}} + 100.6 \frac{\text{kg MeOH}}{\text{day}} = 135.7 \frac{\text{kg MeOH}}{\text{day}} \quad (179)$$

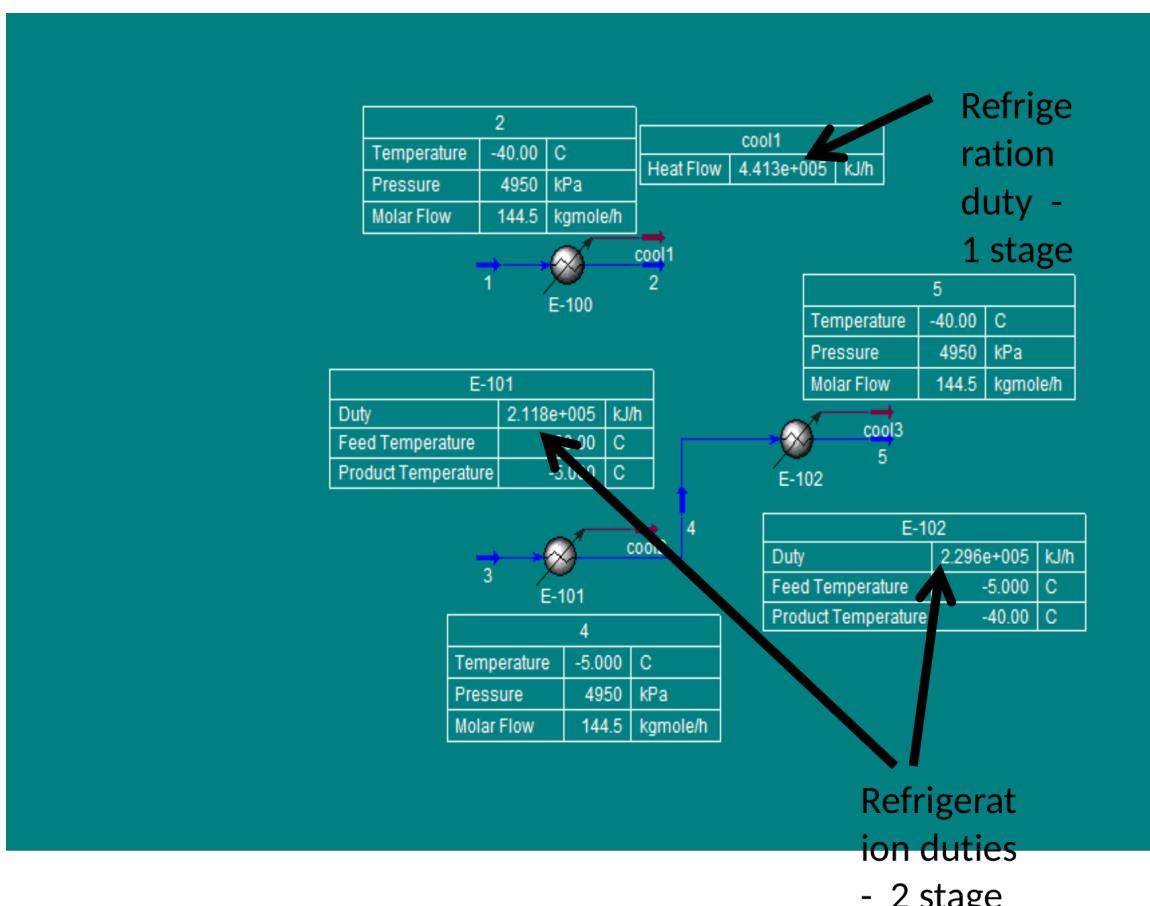
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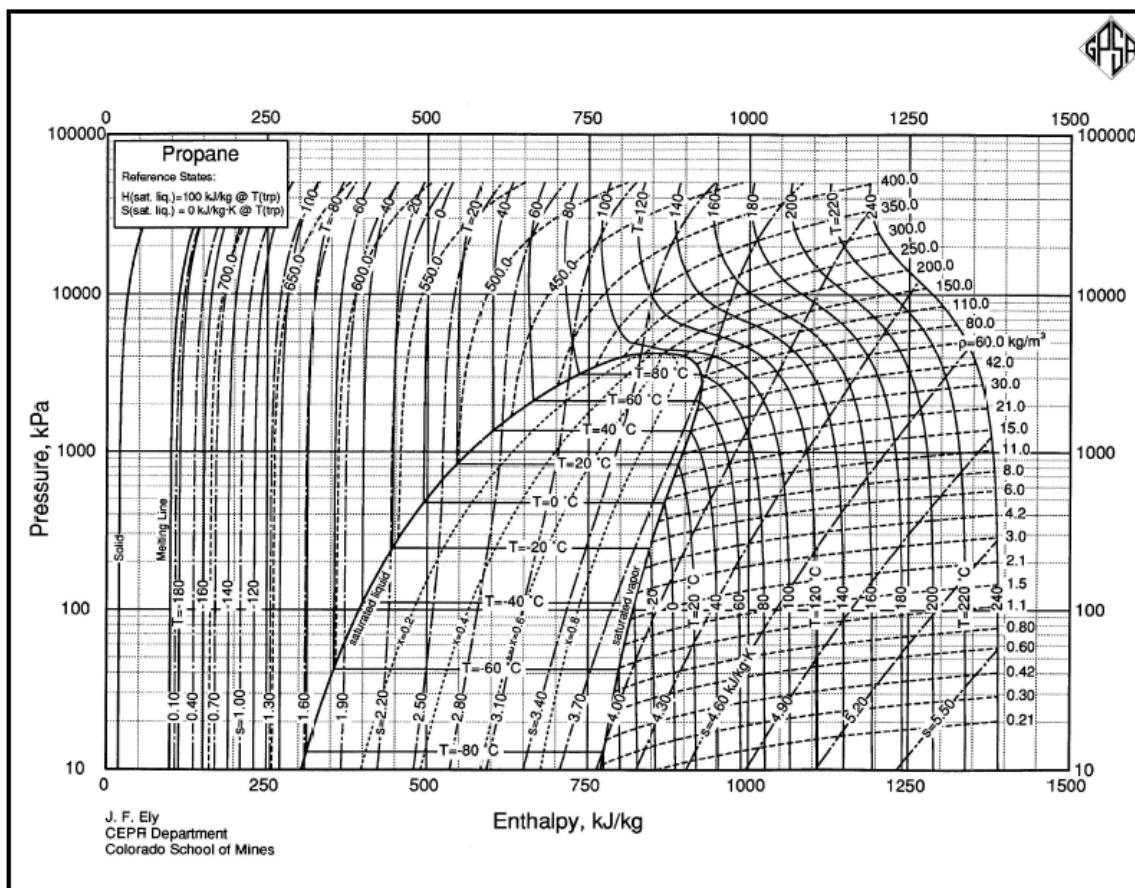
Q.35

Question 35

Use the attached Mollier chart to compare the compressor power requirements and condenser duty of one and two stage refrigeration processes. The refrigerant is propane. The cooling duties are shown in the attached UNISIM output. Use the following assumptions:

- 5 °C approach between refrigerant and process gas outlet
- 5 °C approach to the condensing medium which is air at 20 °C.
- Polytropic efficiency of propane compressor is 80 %.





Solution:

One-stage refrigeration

1. Looking at the Mollier chart, it is required to compress the propane to 1000 kPa because at this pressure the propane condenses at a temperature of 25 °C giving us the required 5 °C temperature approach to the air which is at 20 °C.
2. The propane is expanded to 100 kPa as, at this pressure, the propane will evaporate at 45 °C which gives us the required 5 °C temperature to the process temperature of -40 °C.
3. The Mollier chart gives the compressor suction enthalpy to be 825 KJ kg⁻¹ (point A).
4. Follow the isentropic line from this point to 10 bar which has an enthalpy of 945 KJ kg⁻¹. The isentropic enthalpy change is therefore

$$\Delta H_{isen} = 945 \frac{\text{kJ}}{\text{kg}} - 825 \frac{\text{kJ}}{\text{kg}} = 120 \frac{\text{kJ}}{\text{kg}} \quad (180)$$

5. Correct this value for the compressor efficiency

$$\Delta H_{real} = \frac{\Delta H_{isen}}{E} = \frac{120 \frac{\text{kJ}}{\text{kg}}}{0.8} = 150 \frac{\text{kJ}}{\text{kg}} \quad (181)$$

6. The compressor discharge is located at point B on the Mollier chart and has an enthalpy of

$$825 \frac{\text{kJ}}{\text{kg}} + 150 \frac{\text{kJ}}{\text{kg}} = 975 \frac{\text{kJ}}{\text{kg}} \quad (182)$$

7. The propane will condense at constant pressure to point C with an enthalpy of 570 kJ kg^{-1} .
8. Joule-Thomson (isenthalpic) expansion to 1 bar gives a propane temperature of -45°C .
9. The enthalpy change from point C to point A is the enthalpy that can be extracted from the process and is

$$570 \frac{\text{kJ}}{\text{kg}} + 825 \frac{\text{kJ}}{\text{kg}} = -255 \frac{\text{kJ}}{\text{kg}} \quad (183)$$

10. The total refrigeration duty is $441300 \text{ kJ hr}^{-1}$. Therefore the required propane mass flowrate is

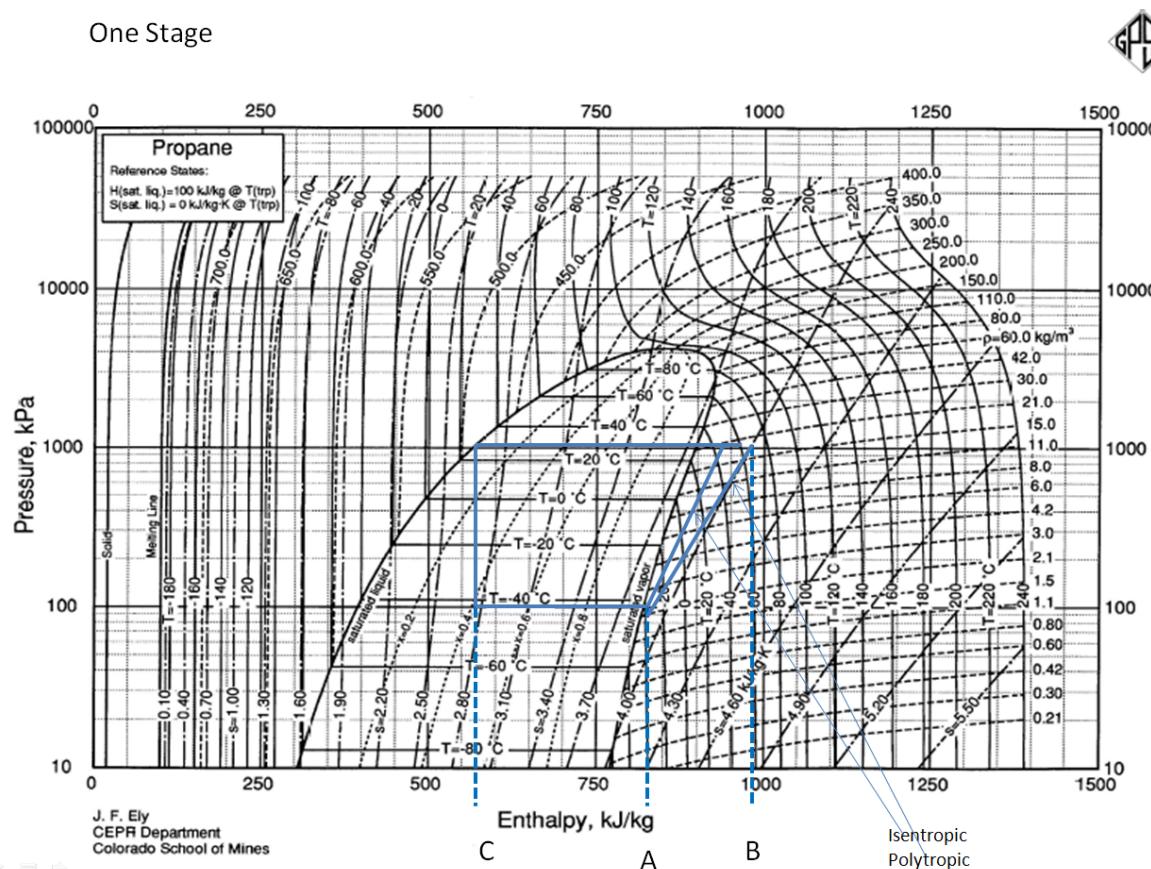
$$m_{\text{propane}} = \frac{441300 \frac{\text{kJ}}{\text{hr}}}{255 \frac{\text{kJ}}{\text{kg}}} = 1730 \frac{\text{kg}}{\text{hr}} \quad (184)$$

11. The compressor duty is equal to propane flowrate multiplied by the compressor duty

$$W_{\text{comp}} = m_{\text{propane}} \Delta H_{\text{real}} = 1730 \frac{\text{kg}}{\text{hr}} \cdot 150 \frac{\text{kJ}}{\text{kg}} = 260000 \frac{\text{kJ}}{\text{hr}} \quad (185)$$

12. The condenser duty is equal to the enthalpy change across the condenser multiplied by the propane flowrate

$$W_{\text{cond}} = \left(975 \frac{\text{kJ}}{\text{kg}} - 570 \frac{\text{kJ}}{\text{kg}} \right) \cdot 1730 \frac{\text{kg}}{\text{hr}} = 700000 \frac{\text{kJ}}{\text{hr}} \quad (186)$$



Two-stage refrigeration

1. The outlet temperature of the first cooler is -5°C . Hence propane at -10°C is required in order to give a 5°C temperature approach.
2. It is only required to expand the propane to 3.5 bar in order to achieve a temperature of -10°C .
3. The compressor suction enthalpy is 855 kJ kg^{-1} and is shown on the attached Mollier chart as point D.
4. As before, following the isentropic line to 10 bar and then correcting for the compressor efficiency gives a compressor output enthalpy of 940 kJ kg^{-1} (point E)
5. The enthalpy change over the compressor is therefore

$$\Delta H_{real} = 940 \frac{\text{kJ}}{\text{kg}} - 855 \frac{\text{kJ}}{\text{kg}} = 85 \frac{\text{kJ}}{\text{kg}} \quad (187)$$

6. The propane condenses at constant pressure to the same point as before at 570 kJ kg^{-1} (point F).
7. Joule Thomson expansion to 3.5 bar gives a propane temperature of -10°C .
8. Enthalpy change from point F to point D

$$570 \frac{\text{kJ}}{\text{kg}} - 855 \frac{\text{kJ}}{\text{kg}} = -285 \frac{\text{kJ}}{\text{kg}} \quad (188)$$

gives the enthalpy that is extracted from the process

9. The required refrigeration for this cooling stage is $211800 \text{ kJ hr}^{-1}$. The required propane flowrate is therefore

$$m_{\text{propane}} = \frac{211800 \frac{\text{kJ}}{\text{hr}}}{285 \frac{\text{kJ}}{\text{kg}}} = 743 \frac{\text{kg}}{\text{hr}} \quad (189)$$

10. The compressor duty is equal to propane flowrate multiplied by the compressor duty

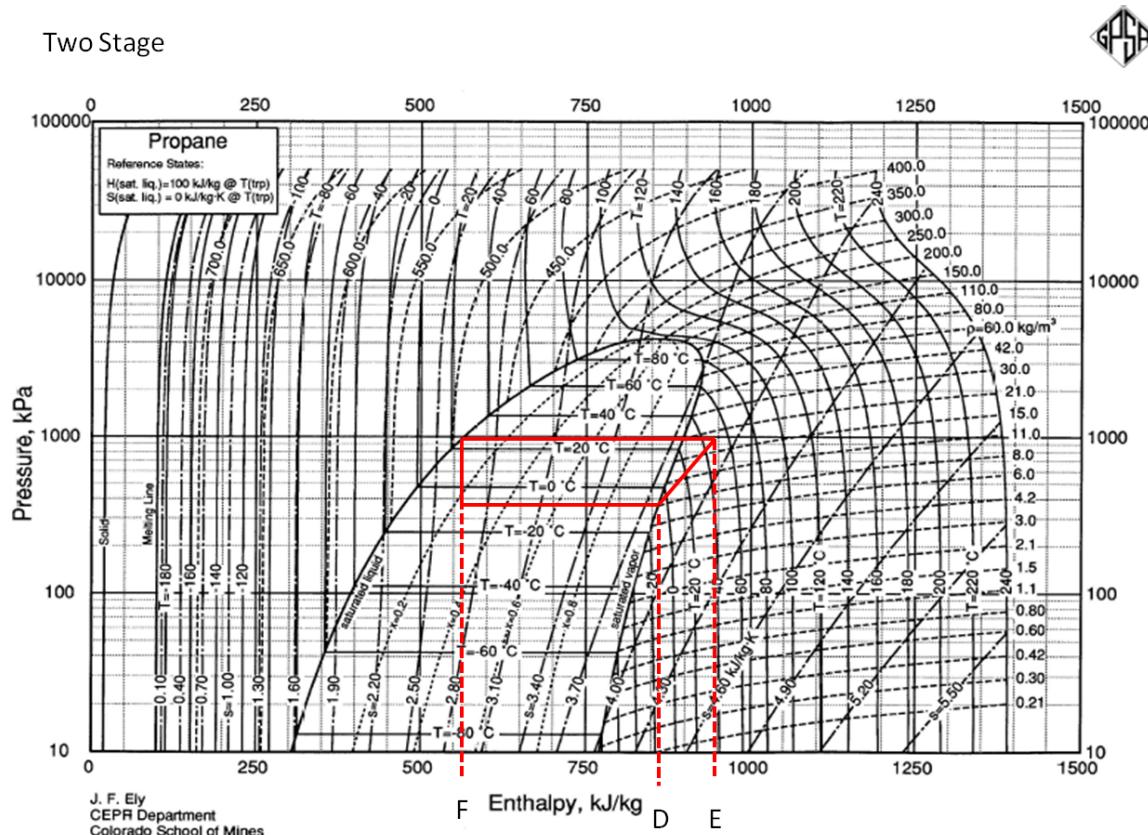
$$W_{\text{comp}} = m_{\text{propane}} \Delta H_{\text{real}} = 743 \frac{\text{kg}}{\text{hr}} \cdot 85 \frac{\text{kJ}}{\text{kg}} = 63100 \frac{\text{kJ}}{\text{hr}} \quad (190)$$

11. The condenser duty is equal to the enthalpy change across the condenser multiplied by the propane flowrate

$$W_{\text{cond}} = \left(940 \frac{\text{kJ}}{\text{kg}} - 570 \frac{\text{kJ}}{\text{kg}} \right) \cdot 743 \frac{\text{kg}}{\text{hr}} = 275000 \frac{\text{kJ}}{\text{hr}} \quad (191)$$

12. For the second cooler, the propane cycle is exactly the same as the previous one-stage option but in this case the refrigeration duty is only $229600 \text{ kJ hr}^{-1}$.

13. The previous propane compressor and condenser duty will reduce by the ratio of the chiller duty (50%). The new duty of the propane compressor becomes $135000 \text{ kJ hr}^{-1}$ and the condenser duty becomes $364000 \text{ kJ hr}^{-1}$.



If we compare the compressor power for the 1 and 2-stage refrigeration options we find that the 2-stage compressor duty

$$W_{\text{comp},2\text{-stage}} = 63100 \frac{\text{kJ}}{\text{hr}} + 135100 \frac{\text{kJ}}{\text{hr}} = 198100 \frac{\text{kJ}}{\text{hr}} \quad (192)$$

is 24% lower than the 1-stage duty

$$W_{comp,1\text{-stage}} = 260000 \frac{\text{kJ}}{\text{hr}} \quad (193)$$

[Question end]

Q.36 Question 36

Past exam question

- a) A process gas is to be cooled from 20°C to -35°C using a refrigeration plant. The refrigerant is propane. Using the attached Mollier chart (see Fig. 6) for propane estimate the compressor energy savings between a one and a two stage refrigeration system. Assume air is available at 25°C for condensation. State any other assumptions. Enclose your Mollier chart including your student number with the returned exam booklet.

[14 marks]

Solution:

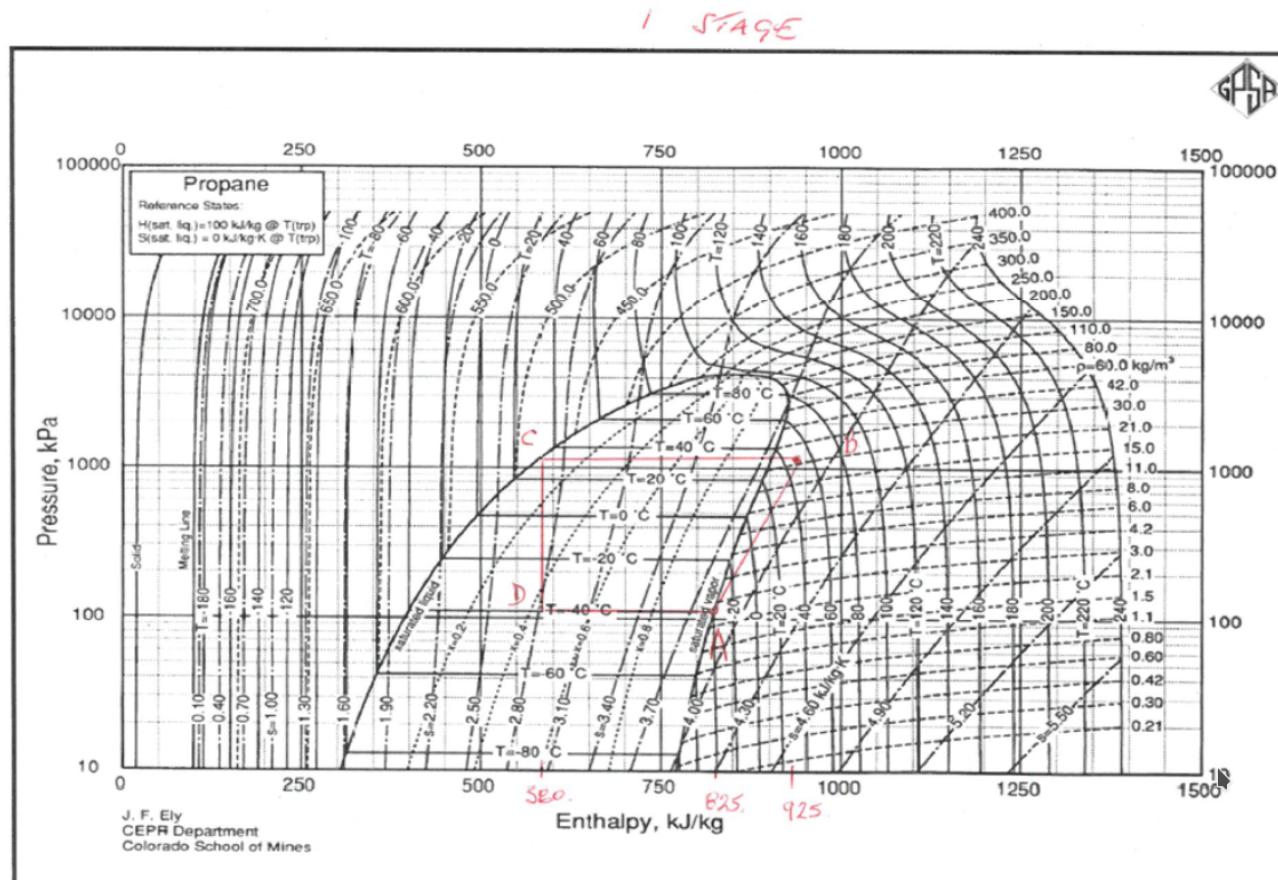
Assume a 5°C temperature approach for condensation and evaporation. That means that the propane will have to be compressed to a pressure that corresponds to $25 + 5 = 30^\circ\text{C}$. From the Mollier chart that is around 1200 kPa. To cool the process to -35°C requires an evaporator pressure of 120 kPa (-40°C). The Mollier chart can now be used to complete the one stage refrigeration system⁴.

[4/14] A to B compression - isentropic enthalpy change = $925 - 825 = 100 \text{ kJ kg}^{-1}$.

[1/14] D to A evaporation duty = $825 - 580 = 245 \text{ kJ kg}^{-1}$.

Hence for 1 kJ s^{-1} of heat removed the propane circulation rate will be $1/245 = 0.0041 \text{ kg s}^{-1}$

[1/14] Hence compressor power = $0.0041 \times 100 = 0.41 \text{ kJ s}^{-1}$.



Compare to 2 stage refrigeration.

Assume first stage cools process from 20 to -7.5°C .

To cool the process to -7.5°C requires evaporator pressure of 350 kPa (-12.5°C). The Mollier chart can now be used to complete the first stage of the two stage refrigeration system^v.

[1/14] A to B compression – isentropic enthalpy change = $925 - 850 = 75 \text{ kJ kg}^{-1}$ ✓.

[1/14] D to A evaporation duty = $850 - 590 = 260 \text{ kJ kg}^{-1}$ ✓.

[1/14] Assume each stage of the two stage process removes half the heat of the one stage system 0.5 kJ s^{-1} of heat removed ✓.

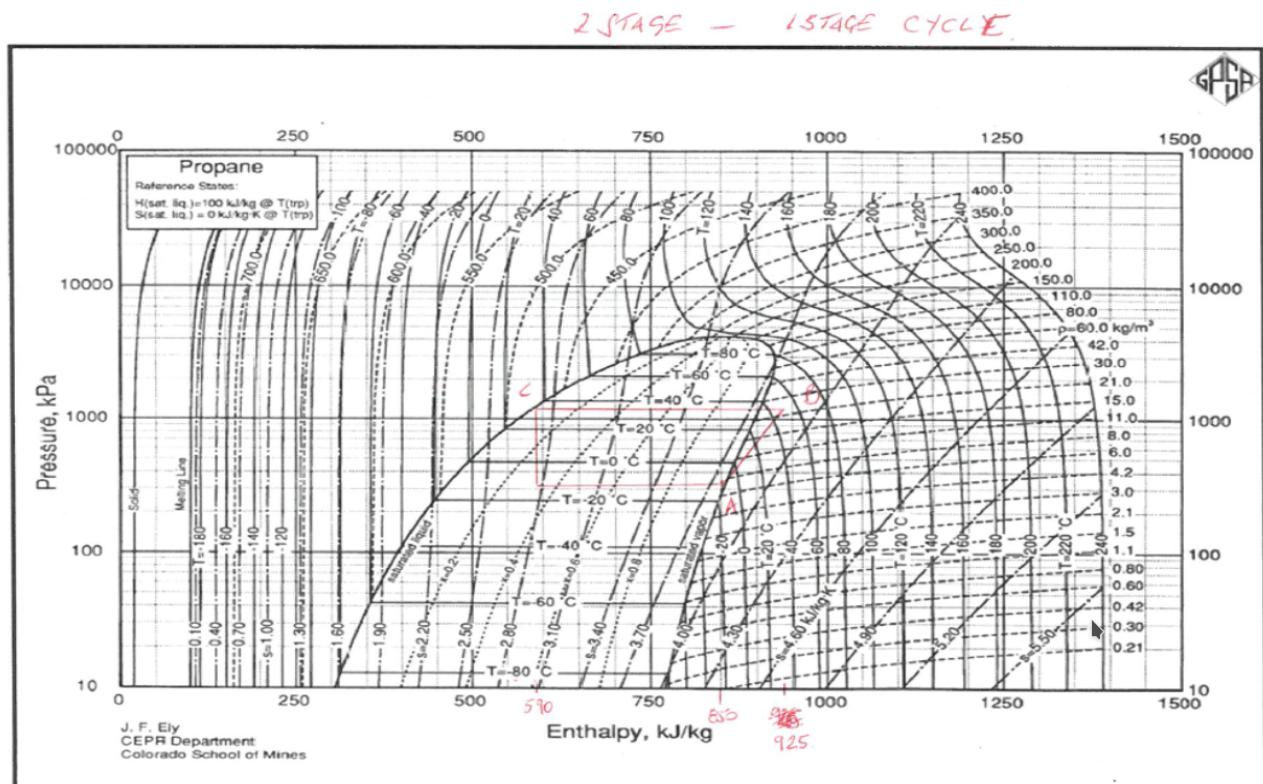
Then the propane circulation rate will be $0.5/260 = 0.0019 \text{ kg s}^{-1}$.

Hence compressor power = $0.0019 \times 75 = 0.1425 \text{ kJ s}^{-1}$.

The second refrigeration stage will be as the one stage process but the heat removed will be half, so the compressor power will also halve to $0.5 \times 0.41 = 0.205 \text{ kJ s}^{-1}$.

[1/14] Total power for two stages = $0.205 + 0.1425 = 0.3475 \text{ kJ s}^{-1}$ ✓.

[1/14] So 2 stages is $0.35/0.41 \times 100 \approx 85\%$ of 1 stage power^v ✓.



- b) State and explain six factors in deciding whether to use a one or two stage refrigeration system. [6 marks]

Solution:

(1 mark each)^v

- (1) Capital cost - two stages more costly - key for economic decision.
- (2) Operating cost - two stage uses less energy - key for economic decision.

- (3) Availability - two stage is more likely to suffer breakdowns as there is more equipment.
- (4) Revenue - key for economic decision.
- (5) Safety - two stage has double the amount of leak paths.
- (6) Environment - two stage has improved carbon footprint.

[Question total: 20 marks]

If you use this graph, you must attach it to your exam booklet using the provided tag.
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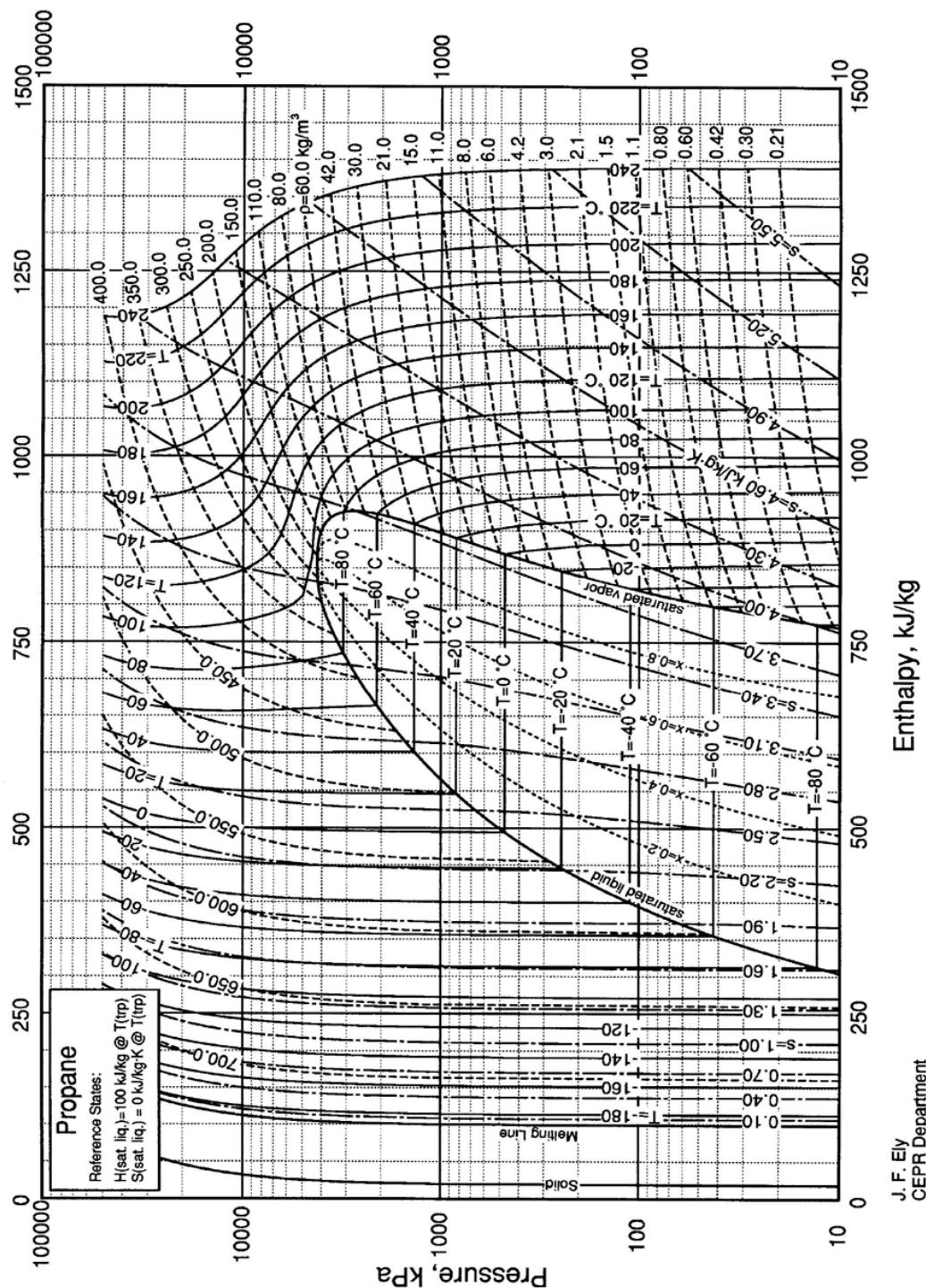


Figure 6: A Mollier chart for propane.

Q.37

Question 37**Past exam question**

- a) Describe hydrates and the main conditions necessary for hydrate formation. Why is hydrate formation a concern for transportation and process systems? [4 marks]

Solution:

Hydrates are formed by a combination of small hydrocarbon molecules - methane to butane, H₂S and CO₂ - and water. The solid crystals formed can lead to flow restrictions and blockages. This could result in an unsafe situation and/or a loss of production. Their formation is favoured by high pressure and low temperature.⁴

[4/4]

- b) Hydrate formation can be calculated using K charts, an example of which is given in Fig. 7.

Explain how these charts could be used to calculate the hydrate formation pressure at a given temperature for a known mixture of methane, ethane, propane and nitrogen. No chart is available for nitrogen.

You should note the following relationship for the dewpoint of a hydrocarbon mixture.

$$\sum_{i=1}^{N_s} \left(\frac{y_i}{K_{vs,i}} \right) = 1$$

where y_i is the mole fraction of species i and $K_{vs,i}$ is the equilibrium ratio between the gaseous/vapour fraction and the solid hydrate for each component. [6 marks]

Solution:

The use of the hydrate charts is analogous to the calculation of a dewpoint for a hydrocarbon mixture. At the dewpoint the above criteria is fulfilled.

Procedure:

- (a) Make an initial guess at the hydrate formation pressure.
- (b) For methane, ethane and propane estimate the $K_{vs,i}$ from the respective charts
- (c) The $K_{vs,i}$ for Nitrogen is infinity as it is not a hydrate former.
- (d) Calculate $y_i/K_{vs,i}$ for each component with the term for Nitrogen equal to zero.
- (e) Complete the summation, if not equal 1 re-guess formation pressure and repeat. ⁶

[6/6]

- c) To prevent hydrate formation a 2366 Sm³ h⁻¹ gas stream is to be dehydrated using a TEG contactor. The gas enters the dehydrator saturated at 80 bar and 30°C and lean TEG is available at 99% purity. Calculate the exit gas water concentration and the flowrate of TEG required assuming 0.04 m³ per kg water removed is required. A safety factor of 10°C is also required.

Equilibrium data are available in Fig. 8 and 9.

[8 marks]

Solution:

From Fig. 8, saturated gas at 80 bar (8000 kPa) and 30°C has an equilibrium concentration of 550 mg Sm⁻³.³

The glycol is available at 99% purity which, according to Fig. 9, has an equilibrium dewpoint of -18°C₂. Unfortunately a safety factor of 10°C is needed, which implies a equilibrium of

[3/8]

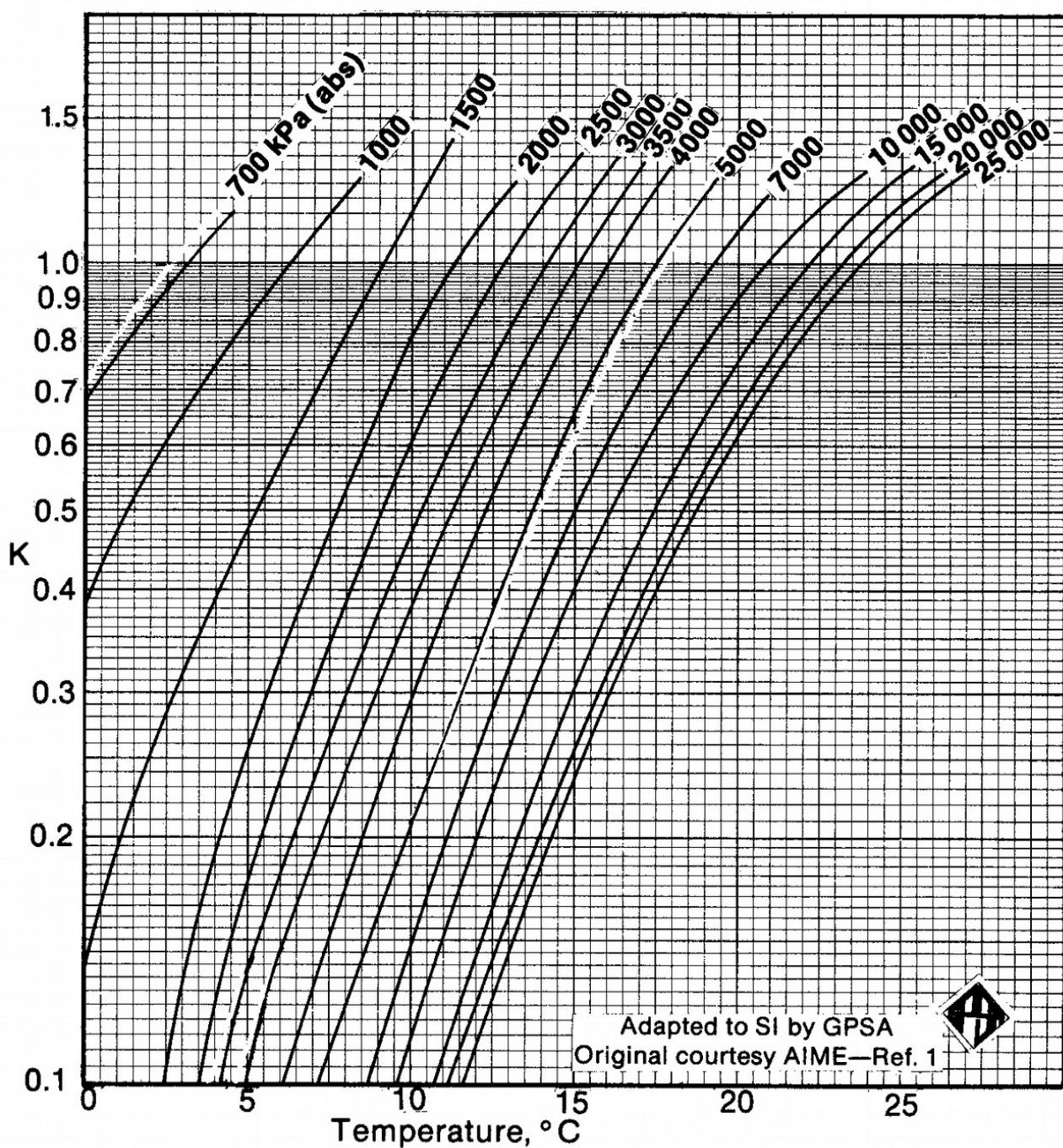


Figure 7: Hydrate K-value chart for ethane.

[2/8] -8°C is used for the calculations. ✓

[1/8] Assuming little/no pressure drop in the contactor, the equilibrium wet gas concentration at 80 bar and -8°C is 55 mg Sm^{-3} according to Fig. 8.1. ✓

The total water removal rate is $2366(550 - 55) = 1171170 \text{ mg h}^{-1}$, or 1.17 kg h^{-1} thus $0.04 \times 1.17 = 0.0468 \text{ m}^3 \text{ h}^{-1} = 1.3 \times 10^{-5} \text{ m}^3 \text{ s}^{-1}$ TEG is required.

- d) What TEG purity would be required to dehydrate the gas to 20 mg Sm^{-3} ? [2 marks]
Solution:

The required dew point is -22.5°C at 8000 kPa. With a safety margin a dewpoint of

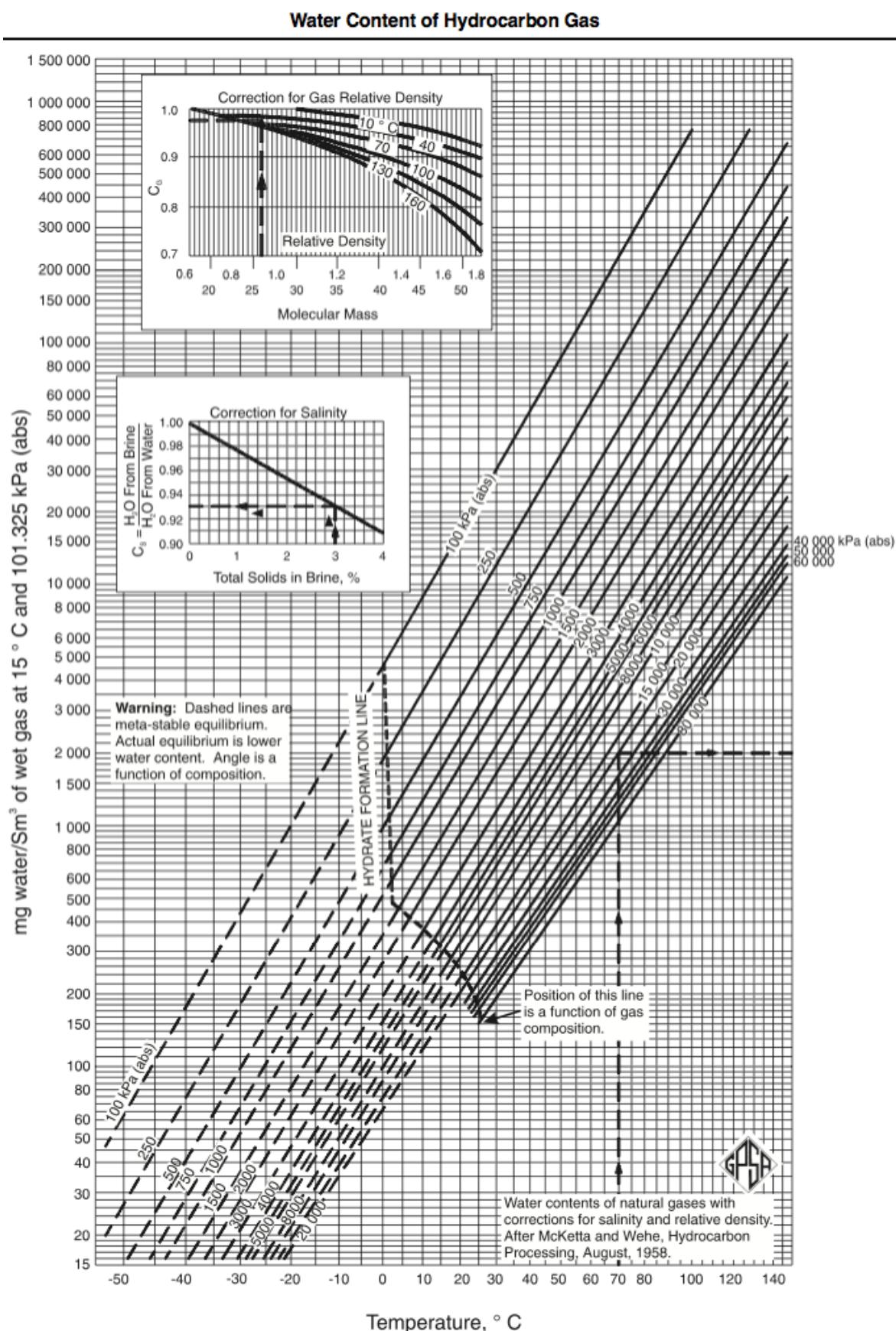


Figure 8: Equilibrium chart for wet natural gas including corrections for water salinity and relative molecular weight/density.

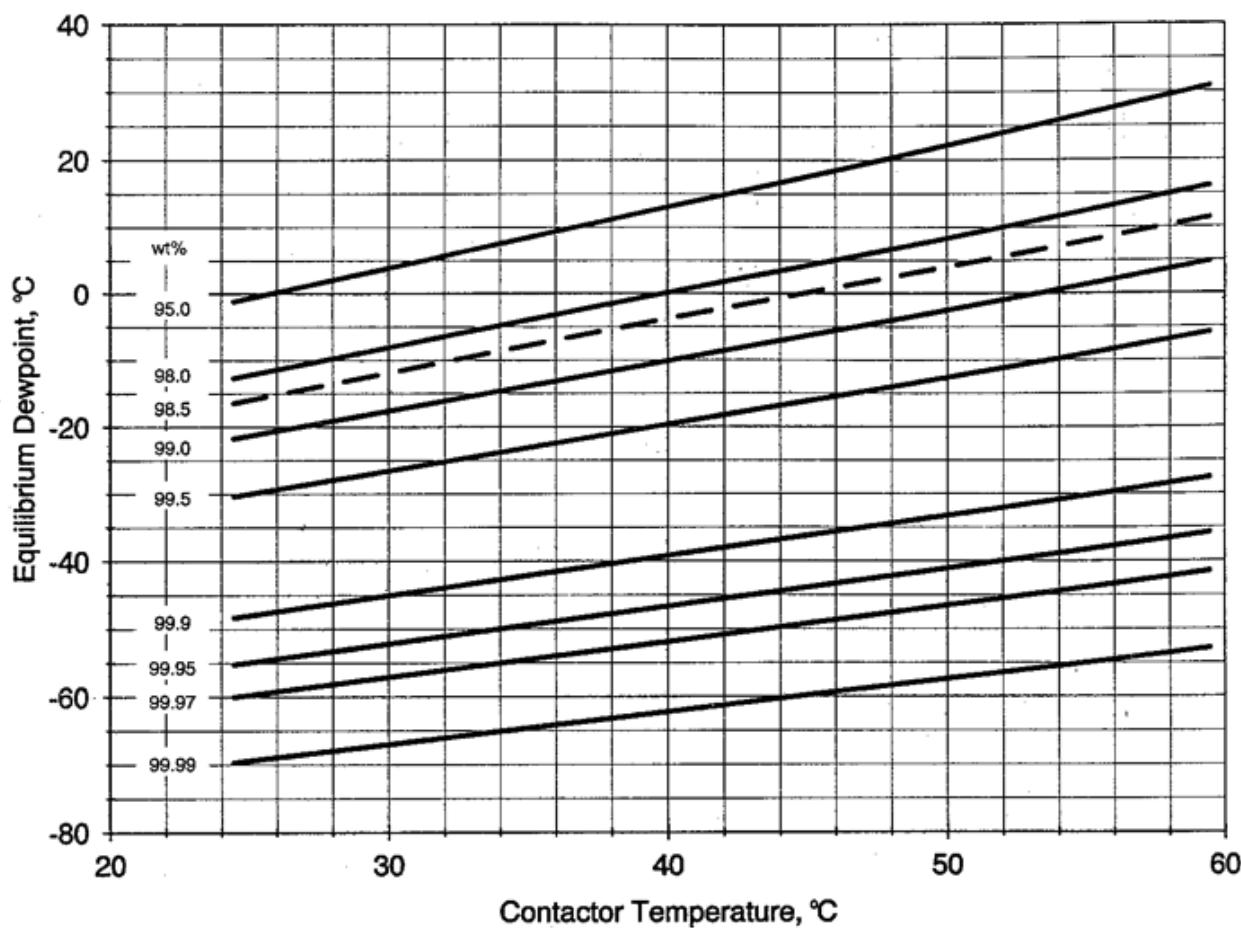


Figure 9: Equilibrium correction chart for wet natural gas over TEG solutions.

[1/2] -32.5°C is required over glycol^v. At a contactor temperature of 30°C , this translates to a
[1/2] $>99.5\%$ but $<99.9\%$ ^v.

[Question total: 20 marks]

Q.38 Question 38 6: Gas treatment

- a) Water is to be removed from a process gas prior to pipeline entry. The pipeline entry specification is 20 mg water/ Sm^3 gas. A glycol absorption unit is available with a lean glycol content of 1.5 wt% water. The absorber pressure and temperature is 50 bara and 35°C , respectively. Will the glycol absorber be capable of achieving the pipeline entry specification? Use the enclosed glycol equilibrium (Fig. 10) and water saturation (Fig. 11) charts and explain your reasoning. **[8 marks]**

Solution:

First, it should be noted that we're discussing if a separation is possible, thus no safety factor should be used as that is just an estimate of the optimal design case (and will be adjusted through the detailed design), not the maximum achievable performance^v.

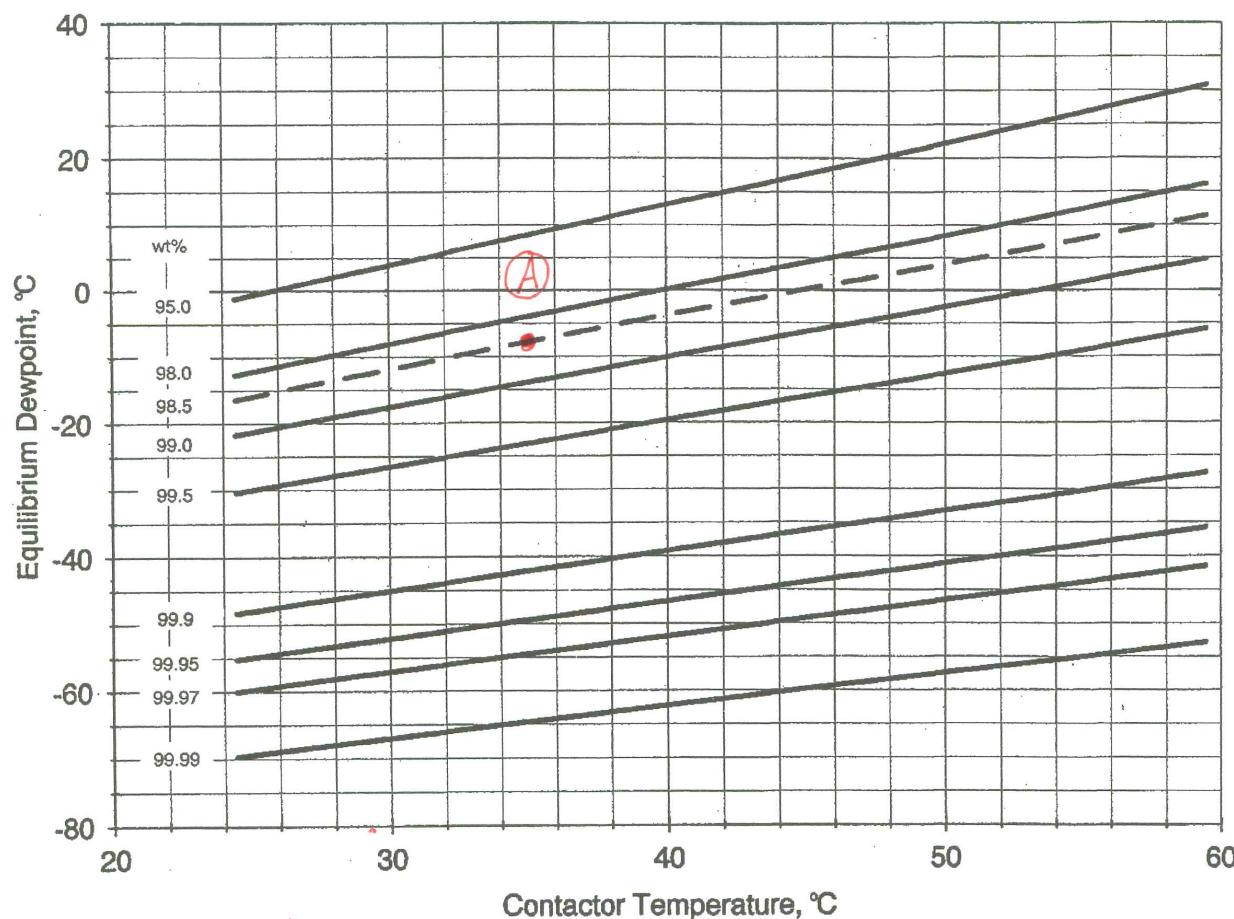
The gas leaving the top of the glycol absorber requires to have a water content of 20mg water/ Sm^3 gas. Water is removed from the gas by absorption into the counter flowing glycol hence the lowest possible water content would occur if the glycol and the water reached equilibrium.^v

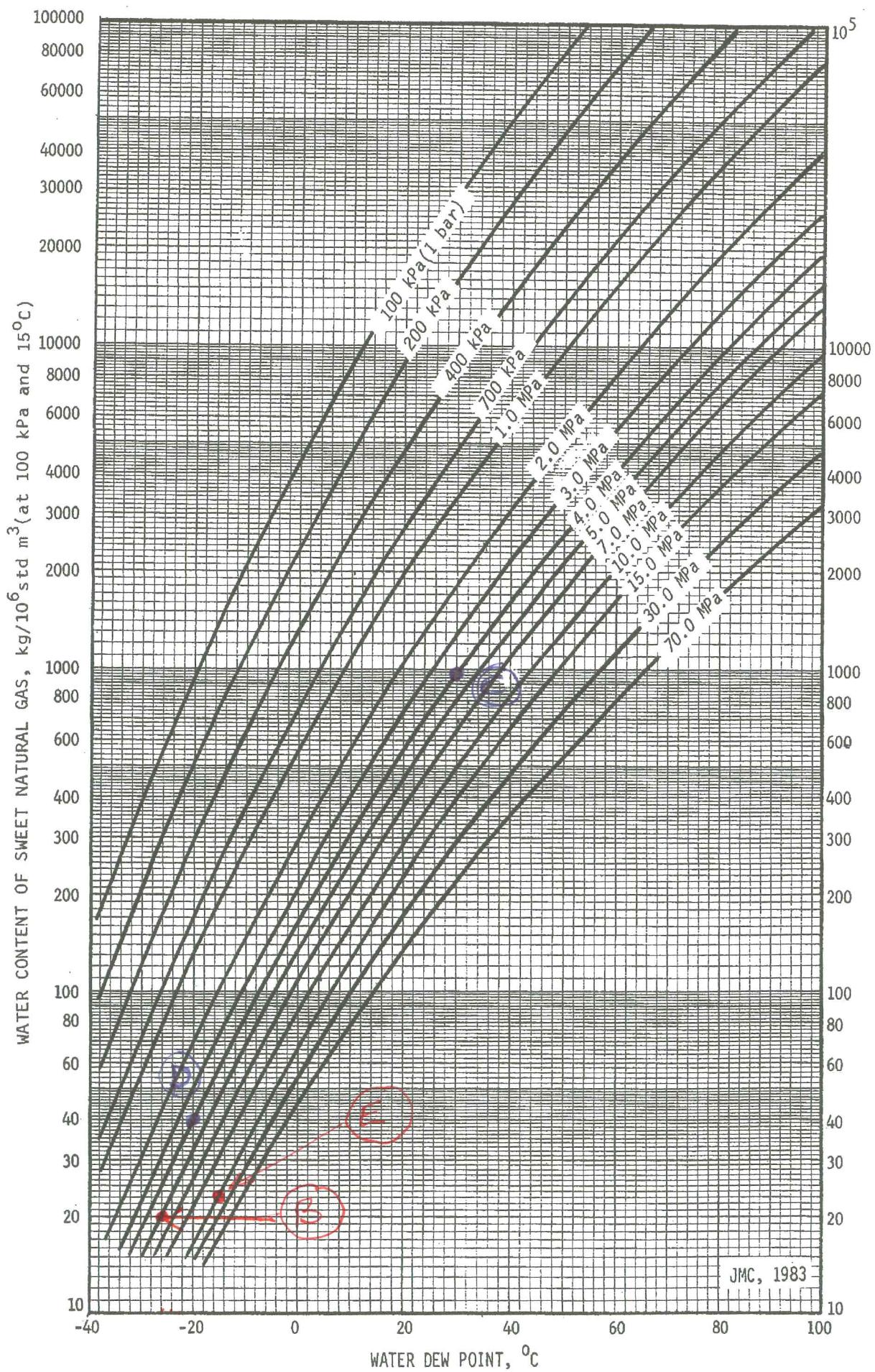
The equilibrium level has to be determined for the contactor conditions. The Glycol water content is 1.5 wt% water-98.5 wt% glycol. From equilibrium chart at 35°C the equilibrium dew point is indicated by A - a dewpoint temperature of -8°C which corresponds to 80 mg Sm⁻³.[✓]

[3/8] Comparing the dew point temperature with the dew point temperature of a gas with 20mg water/sm³. Use water content chart - at contactor pressure (50 bara) the dew point temperature is -26°C (point B).[✓] Hence the contactor could not achieve the required pipeline specification of 20mg water/Sm³ gas.[✓] Alternatively we could compare what the dewpoint temperature of -8°C at 50 bara corresponds to 80 mg Sm⁻³, which is again above the specification.

To achieve the required specification, a glycol purity of 99.7% is required.

(allow for inaccuracies with chart reading)





- b) Describe the key aspects of the molecular sieve process used for gas drying. [6 marks]

Solution:

Molecular sieves adsorb water molecules contained within the gas onto a solid surface. The process requires two (or more) vessels with one on-line removing water while the other is being regenerated. The process is normally on a time cycle - typically 4-24 hours. When the bed is taken off-line to be regenerated, the water is removed by heating the sieve to typically 230-320°C, depending on the desiccant used and the performance specification. The bed is heated using regeneration gas. The regeneration gas used to heat the bed is a slipstream of dry process gas or the feed gas. The regeneration gas is returned to the process after it has been cooled and the free water removed. Following the heating cycle the heater is switched off and the bed cooled.

Gas flow during adsorption is typically downflow. This allows higher gas velocities (thus smaller diameter towers) since bed fluidization is avoided. Regeneration gas flow is upflow during the heating cycle. In that way, any residual water left on the desiccant will be at the top of the bed and will not affect the effluent dewpoint when adsorption is resumed. In addition, upflow heating helps to strip any contaminants from the top of the bed extending desiccant life. Regeneration gas flow during the cooling cycle may be upflow if the gas is completely free of water. This saves two switching valves per tower which can reduce capital costs. If the cooling gas contains water, cooling flow should be downflow to avoid preloading of the desiccant at the bottom of the bed with water.

Finally, the note of the phenomena of breakthrough as well as the corresponding zones of saturation, mass transfer, and dry bed material are relevant points for discussion.

Key points for student to answer.(1 mark each, any 6 points are accepted)

- (a) Adsorption of water onto a solid surface.
 - (b) Process involves one bed adsorbing whilst other regenerating.
 - (c) Process on a time cycle
 - (d) Regeneration requires heated slip stream of gas
 - (e) Bed requires to be cooled before being brought in-line
 - (f) Bed consists of a saturated/equilibrium zone, a mass transfer zone, and a depleted/fresh zone ready for transfer.
 - (g) Breakthrough of the water through the bed is designed to occur towards the end of design life (i.e. once the effectiveness of the adsorbent has degraded).
- c) A molecular sieve dehydration unit is operating at 40 bara and 30°C. The unit is required to produce a product natural gas with a water dewpoint of minus 20°C. The gas entering the unit is water saturated and the flowrate is $2 \times 10^6 \text{ sm}^3/\text{day}$. How much water is the unit required to remove? You will need the data in Fig. 11. [3 marks]

Solution:

Use the water saturation chart. Gas enters mol sieve at point C ($960\text{kg}/10^6 \text{ sm}^3$) and leaves at point D ($40\text{kg}/10^6 \text{ sm}^3$). Hence amount of water removed is, $960 - 40 = 920 \text{ kg}/10^6 \text{ sm}^3$. Thus amount of water removed = $2 \times 10^6 \times 920/10^6 = 1840 \text{ kg day}^{-1} = 76.7 \text{ kg hr}^{-1} = 0.0213 \text{ kg s}^{-1}$.

(allow for inaccuracies with chart reading)

- d) The export gas from the above molecular sieve later enters a cold plant operating at 150 bara and minus 15°C. Comment on the suitability of the molecular sieve treated gas for the cold plant operation. [3 marks]

Solution:

The conditions of the cold plant are indicated by point E (23 mg Sm^{-3}). The operating conditions of the cold plant are below the dew point temperature of the gas leaving the mol sieve. Hence water will condense within the cold plant. This would be unacceptable as the plant would be at risk from ice and hydrate formation.

(allow for inaccuracies with chart reading)

[Question total: 20 marks]

If you use this graph, you must attach it to your exam booklet using the provided tag.
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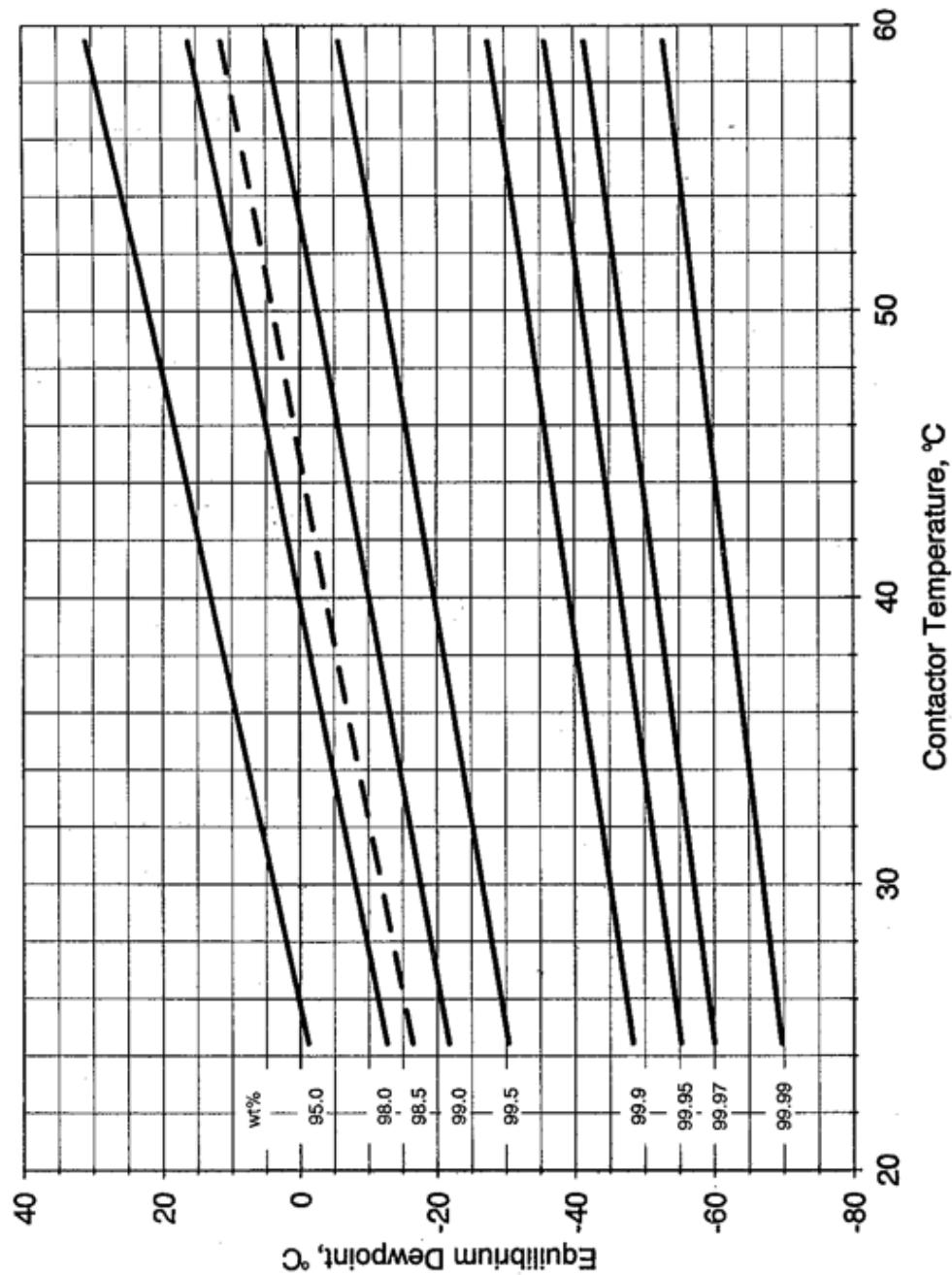


Figure 10: A TEG equilibrium chart.

If you use this graph, you must attach it to your exam booklet using the provided tag.
 Student ID:

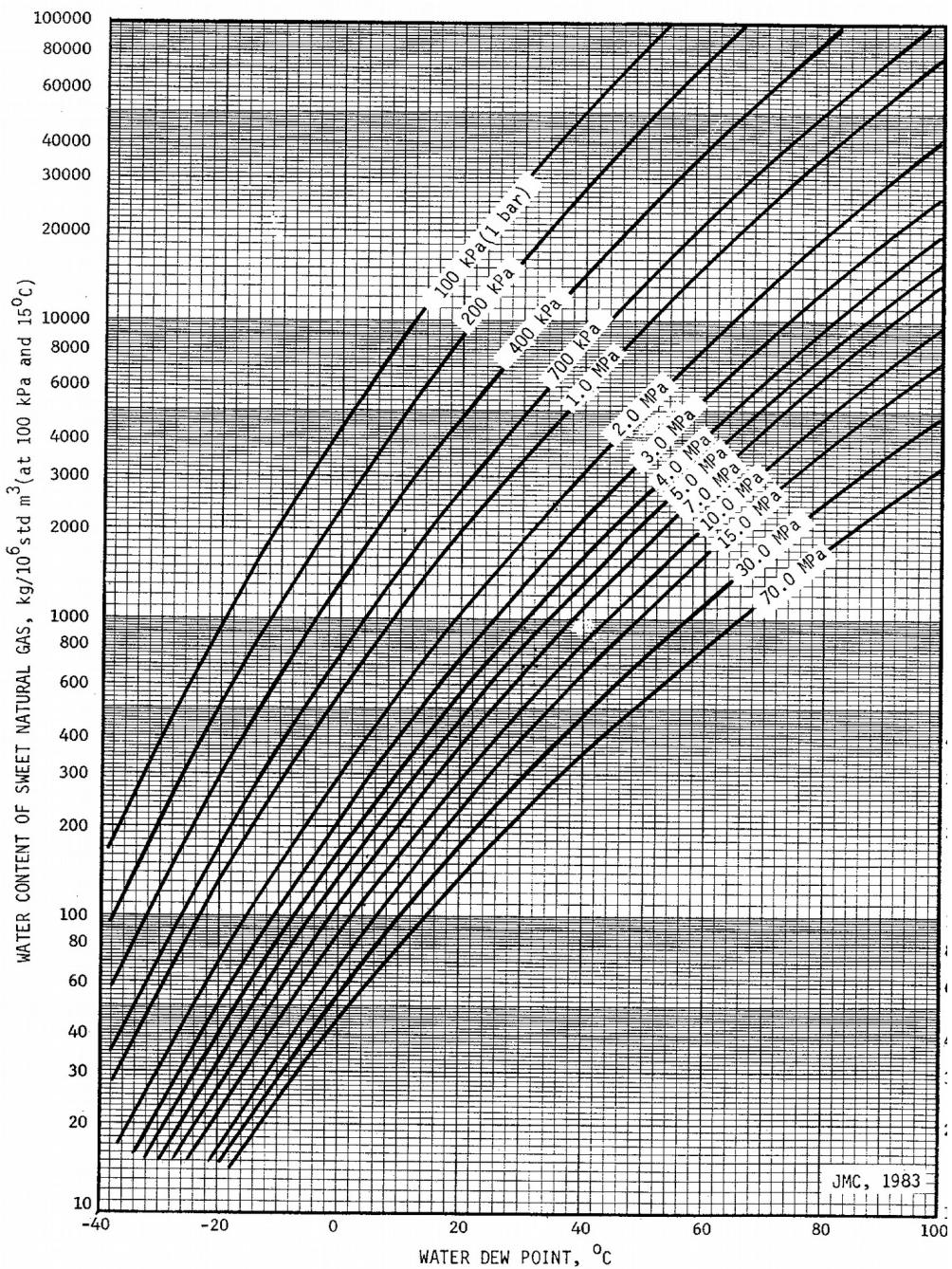


Figure 11: Sweet-gas/water dew point graph.

7 7: Flow assurance (Multi-phase flow)

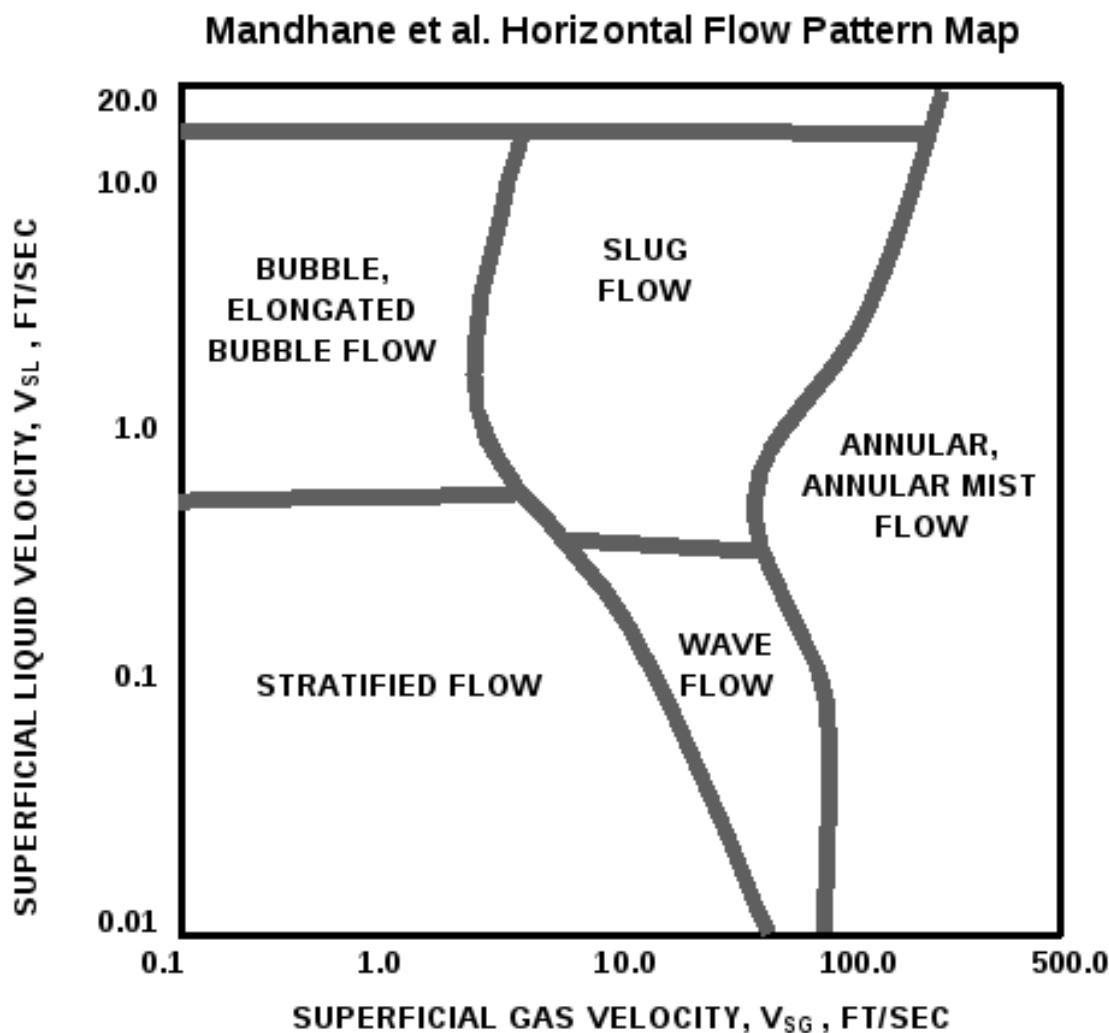
Q.39 Question 39

Two phase flow is present in a pipeline with the following conditions:

- Pipeline diameter = 0.381 m
- Gas volume flow = $0.835 \text{ m}^3 \text{ s}^{-1}$

- Liquid volume flow = $0.0369 \text{ m}^3 \text{ s}^{-1}$

Using the attached Mandhane flow map, predict the expected flow pattern. [slug]



Solution:

The area of the pipe is

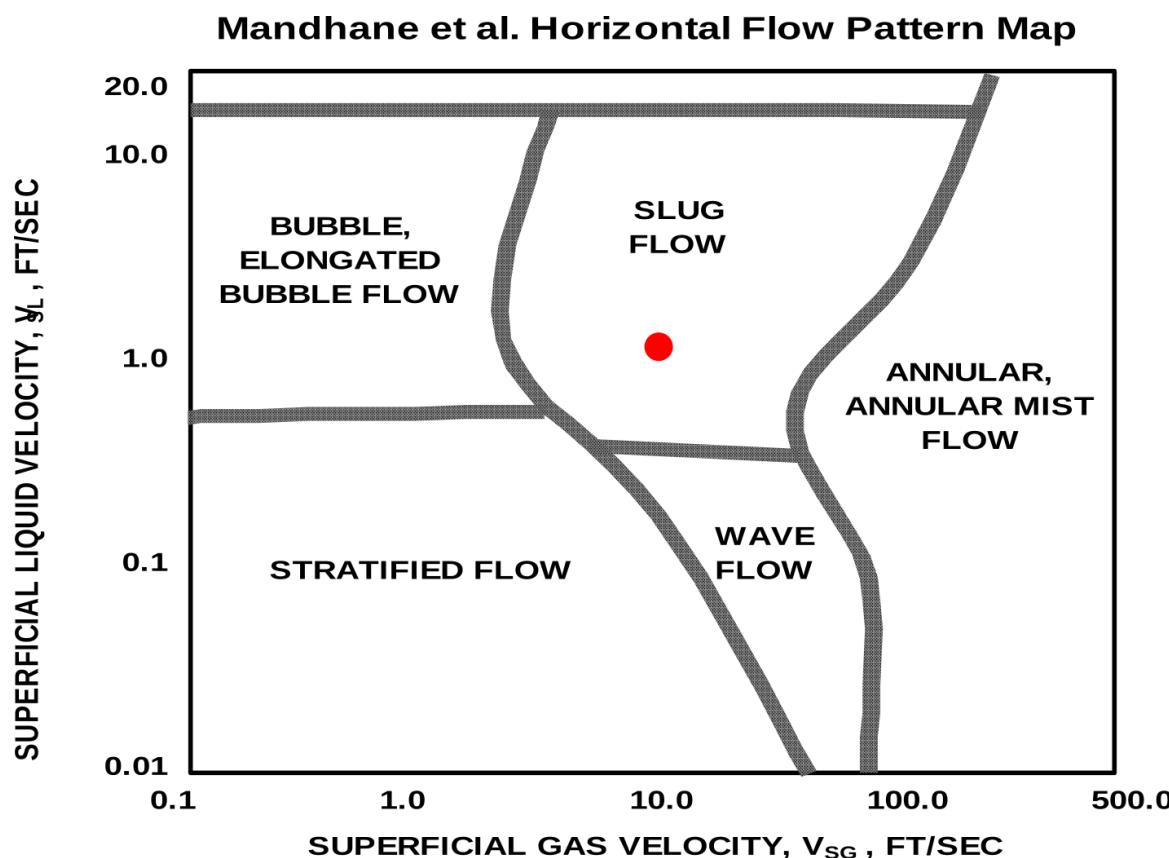
$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot 0.381 \text{ m}}{4} = 0.114 \text{ m}^2 \quad (194)$$

This is used to calculate the superficial liquid and gas velocities

$$v_{SL} = \frac{Q_L}{A} = \frac{0.0369 \text{ m}^3 \text{ s}^{-1}}{0.114 \text{ m}^2} = 0.323 \text{ m s}^{-1} = 1.06 \text{ ft s}^{-1} \quad (195)$$

$$v_{SG} = \frac{Q_G}{A} = \frac{0.835 \text{ m}^3 \text{ s}^{-1}}{0.114 \text{ m}^2} = 7.32 \text{ m s}^{-1} = 24.0 \text{ ft s}^{-1} \quad (196)$$

This point is shown on the attached Mandhane flow map which predicts slug flow.

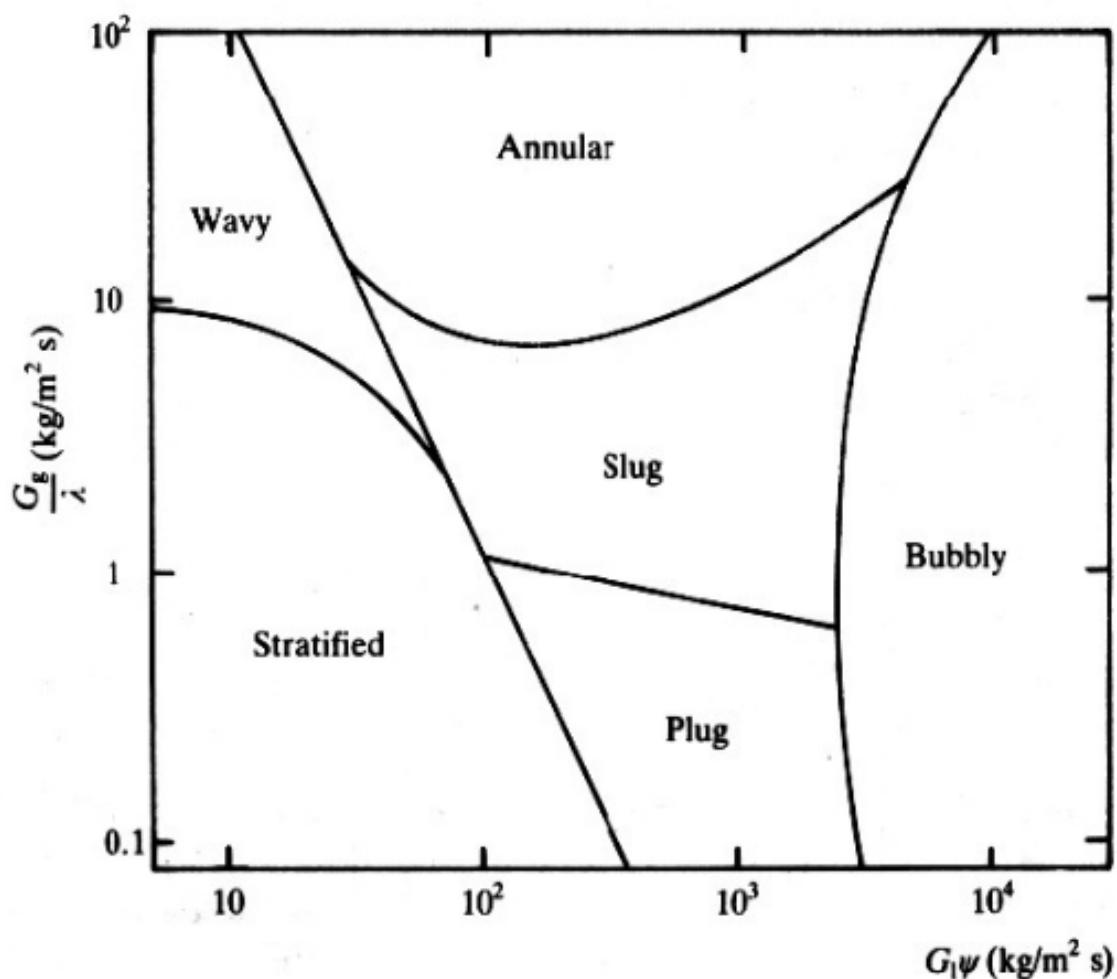


[Question end]

Q.40 Question 40

Using the Baker flow pattern map evaluate the most likely flow pattern occurring in a 2.54 cm i.d. horizontal pipe when the system pressure is 30 bar, 70 bar, and 170 bar, the gas mass quality x is 1 %, 10 %, and 50 %, and the mass velocity G is $500 \text{ kg m}^{-2} \text{ s}^{-1}$, and $2000 \text{ kg m}^{-2} \text{ s}^{-1}$. Fluid properties are:

- Liquid density = $817, 741, 556 \text{ kg m}^{-3}$ for 30, 70, 170 bara respectively
- Gas density = $15, 35.6, 119.3 \text{ kg m}^{-3}$ for 30, 70, 170 bara respectively
- Liquid viscosity = 0.00014 Pa s
- Gas viscosity = 0.0000169 Pa s
- Water surface tension = $0.029835, 0.017633, 0.0032375 \text{ N m}^{-1}$ for 30, 70, 170 bara respectively



Solution:

Calculate the two terms λ and ψ

$$\lambda = \sqrt{\rho'_L \rho'_G} \quad (197)$$

$$\psi = \frac{1}{\sigma'} \left(\frac{\mu'_L}{(\rho'_L)^2} \right)^{\frac{1}{3}} \quad (198)$$

where σ is the surface tension, $\rho'_G = \rho_G / \rho_{\text{atm}}$ is the reduced gas density, $\rho'_L = \rho_L / \rho_{H_2O}$ is the reduced liquid density, $\mu'_L = \mu_L / \mu_{H_2O}$ is the reduced liquid viscosity, and $\sigma'_L = \sigma_L / \sigma_{H_2O}$ is the reduced liquid surface tension.

Note that $\rho_{\text{atm}} = 1.2 \text{ kg m}^{-3}$, $\rho_{H_2O} = 1000 \text{ kg m}^{-3}$, $\mu_{H_2O} = 0.001 \text{ Pa s}$, and $\sigma_{H_2O} = 0.073 \text{ N m}^{-1}$.

Use the Baker flow map using Gx/λ on the y-axis and $G(1 - x)\psi$ on the x-axis.

| p | 30 | 70 | 170 | |
|------------|-------------------|----------------------------|-------------------|-------------------|
| σ_l | 0.029835 | 0.017633 | 0.0032375 | N/M |
| ρ_l | 817 | 741 | 556 | kg/m ³ |
| ρ_g | 15 | 36.5 | 119.3 | kg/m ³ |
| λ | 3.19570023625496 | 4.74749934175877 | 7.43476070002705 | |
| σ' | 0.408698630136986 | 0.241547945205479 | 0.044349315068493 | |
| Ψ | 1.45375579458261 | 2.62518869793872 | 17.315667343899 | |
| x | $G(1 - x)\Psi$ | | | |
| 0.01 | 719.609118318393 | 1299.46840547967 | 8571.25533523002 | |
| 0.1 | 654.190107562175 | 1181.33491407243 | 7792.05030475456 | |
| 0.5 | 363.438948645653 | 656.297174484681 | 4328.91683597476 | |
| x | Gx/λ | | | |
| 0.01 | 1.56460231885188 | 1.05318603333343 | 0.672516601641504 | |
| 0.1 | 15.6460231885188 | 10.5318603333343 | 6.72516601641504 | |
| 0.5 | 78.2301159425939 | 52.6593016666716 | 33.6258300820752 | |
| x | Pattern | | | |
| 0.01 | Slug | Slug \rightarrow Bubbly | Bubbly | |
| 0.1 | Annular | Slug \rightarrow annular | Bubbly | |
| 0.5 | Annular | Annular | Annular | |

[Question end]

Q.41 Question 41

- Repeat the Eaton hold-up calculation in the lecture notes but use a 250 mm pipe diameter. All relevant parameters are given below:
 - Pipeline internal diameter $D = 250$ mm
 - Liquid viscosity $\mu_L = 0.02$ Pa s
 - Liquid flowrate $Q_L = 17$ m³ hr⁻¹
 - Vapour flowrate $Q_G = 425$ m³ hr⁻¹
 - Liquid density $\rho_L = 880$ kg m⁻³
 - Gas density $\rho_G = 20.8$ kg m⁻³
 - Surface tension $\sigma = 1.5 \times 10^{-6}$ N m⁻¹
 - Inlet pressure $P_{in} = 2800$ kPa
 - Outlet pressure $P_{out} = 2400$ kPa
 - $P_b = 101.6$ kPa
- The pipeline is 50 km long, calculate the total liquid hold-up in m³.
- The pipeline terminates with a slugcatcher to capture liquids when the pipe is sphered. The liquid capacity of the slugcatcher is 50 m³. How frequently will the line require sphering to remain within the liquid capacity?

Solution:

Calculate the pipe area

$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot (0.25 \text{ m})^2}{4} = 0.049 \text{ m}^2 \quad (199)$$

Calculate the superficial liquid and gas velocities

$$V_{SL} = \frac{Q_L}{A} = \frac{17 \frac{\text{m}^3}{\text{hr}}}{0.049 \text{ m}^2 \cdot 3600 \frac{\text{s}}{\text{hr}}} = 0.096 \text{ m s}^{-1} \quad (200)$$

$$V_{SG} = \frac{Q_G}{A} = \frac{425 \frac{\text{m}^3}{\text{hr}}}{0.049 \text{ m}^2 \cdot 3600 \frac{\text{s}}{\text{hr}}} = 2.41 \text{ m s}^{-1} \quad (201)$$

Calculate dimensionless groups

$$N_{LV} = 0.0565 V_{SL} \left(\frac{\rho_L}{\sigma} \right)^{0.25} = 0.846 \quad (202)$$

$$N_{GV} = 0.0565 V_{SG} \left(\frac{\rho_L}{\sigma} \right)^{0.25} = 21.1 \quad (203)$$

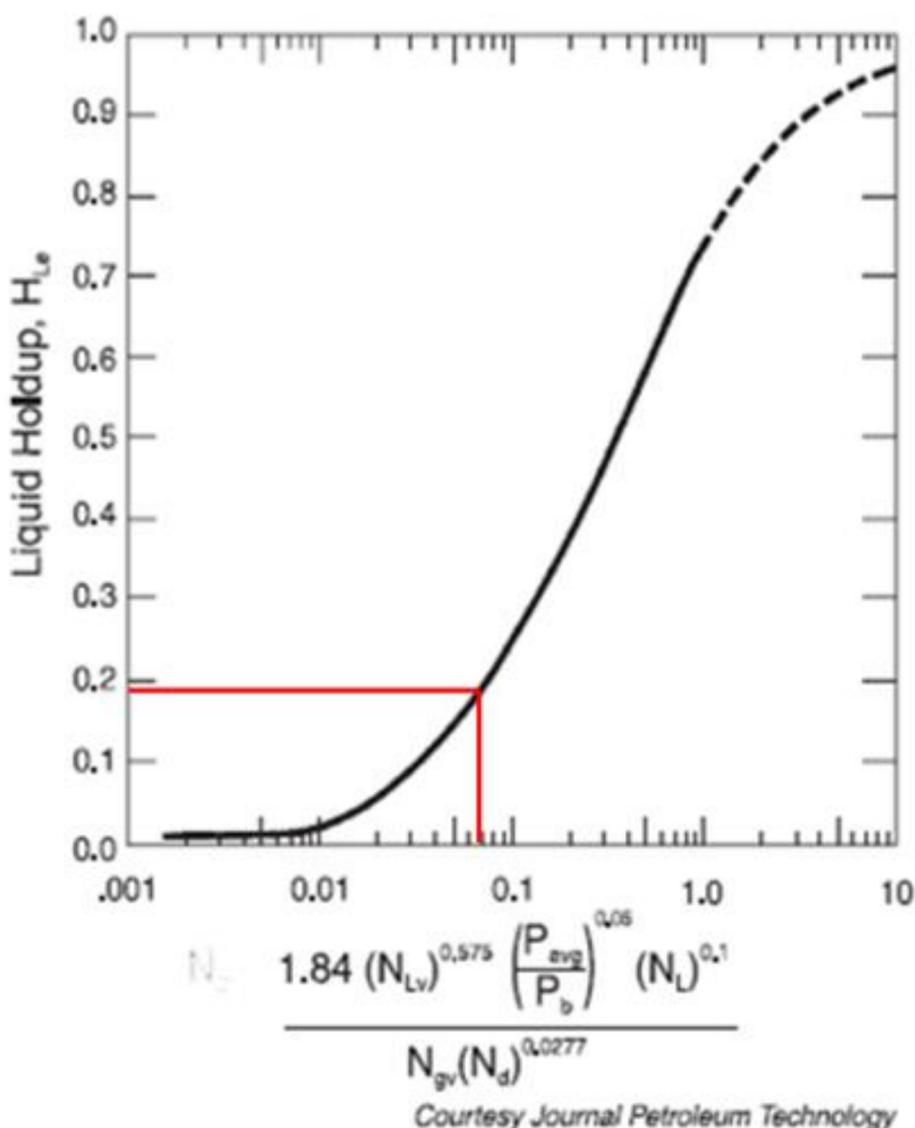
$$N_{GV} = 0.00003134 d \left(\frac{\rho_L}{\sigma} \right)^{0.5} = 190 \quad (204)$$

$$N_{GV} = 0.001769 \mu_L \left(\frac{1}{\rho_L \sigma^3} \right)^{0.25} = 0.15 \quad (205)$$

Calculate Eaton's abscissa

$$N_e = \frac{1.84 (N_{LV})^{0.575} \left(\frac{P_{avg}}{P_b} \right)^{0.05} (N_L)^{0.1}}{N_{GV} (N_d)^{0.0277}} = 0.0665 \quad (206)$$

Use the attached graph to obtain a liquid hold-up fraction of 19 %.



- The pipeline is 50 km long, calculate the total liquid hold-up in m³.

The total volume of the pipe is

$$V_{pipe} = 50000 \text{ m} \cdot 0.049 \text{ m}^3 = 2450 \text{ m}^3 \quad (207)$$

Multiply this with the liquid hold-up fraction to obtain the total liquid hold-up

$$\text{Total liquid hold up} = 0.19 \cdot 2450 \text{ m}^3 = 466 \text{ m}^3 \quad (208)$$

- The pipeline terminates with a slugcatcher to capture liquids when the pipe is spherized. The liquid capacity of the slugcatcher is 50 m³. How frequently will the line require spherizing to remain within the liquid capacity?

The line must be spherized once the total liquid contained in the line is 50 m³.

$$t = \frac{V_{slug\ catcher}}{Q_L} = \frac{50 \text{ m}^3}{17 \frac{\text{m}^3}{\text{hr}}} = 2.96 \text{ hr} \quad (209)$$

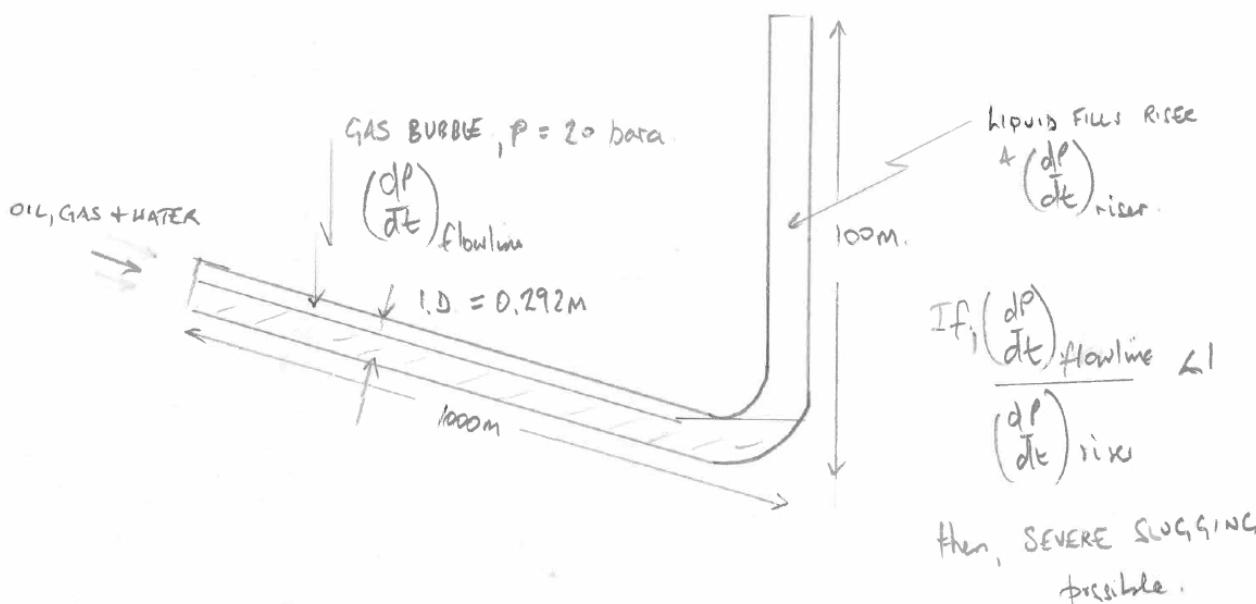
[Question end]

Q.42 Question 42

A downward sloping 1000 m pipeline with an internal diameter of 0.292 m joins a 100 m long vertical pipe (riser) of the same diameter. The pipeline contains oil, gas, and water flowing at an average pressure of 20 bara. The flow rates are:

- Oil = $0.009 \text{ m}^3 \text{ s}^{-1}$
- Water = $0.009 \text{ m}^3 \text{ s}^{-1}$
- Gas = $0.041 \text{ m}^3 \text{ s}^{-1}$

The oil and water densities are 800 and 1000 kg m^{-3} respectively. Determine if the system of pipeline and riser will be prone to severe slugging at the given flow conditions.

**Solution:**

The cross sectional area of the pipeline and riser is

$$A = \frac{\pi D^2}{4} = \frac{\pi (0.292 \text{ m})^2}{4} = 0.067 \text{ m}^2 \quad (210)$$

The system will be prone to severe slugging if

$$\left(\frac{dp}{dt}\right)_{\text{riser}} > \left(\frac{dp}{dt}\right)_{\text{pipeline}} \quad (211)$$

Riser

If the riser is filling with liquid only then the liquid velocity in the riser will be

$$v_{L,\text{riser}} = \frac{Q_L}{A} = \frac{0.018 \frac{\text{m}^3}{\text{s}}}{0.067 \text{ m}^2} = 0.269 \frac{\text{m}}{\text{s}} \quad (212)$$

The liquid mixture density in the riser will be

$$\rho_{mL} = \frac{Q_{oil}}{Q_{oil} + Q_{water}} \rho_{oil} + \frac{Q_{water}}{Q_{oil} + Q_{water}} \rho_{water} = 0.5 \cdot 800 \frac{\text{kg}}{\text{m}^3} + 0.5 \cdot 1000 \frac{\text{kg}}{\text{m}^3} = 900 \frac{\text{kg}}{\text{m}^3} \quad (213)$$

The rate of pressure increase in the riser will be

$$\left(\frac{dp}{dt} \right)_{\text{riser}} = \rho_L g \frac{dh}{dt} = \rho_{mL} g v_{L,\text{riser}} = 900 \frac{\text{kg}}{\text{m}^3} \cdot 9.81 \frac{\text{m}}{\text{s}^2} \cdot 0.269 \frac{\text{m}}{\text{s}} = 0.02373 \frac{\text{bar}}{\text{s}} \quad (214)$$

Pipeline

The no-slip liquid hold up in the pipeline is

$$\lambda_L = \frac{Q_L}{Q_L + Q_G} = \frac{0.018 \frac{\text{m}^3}{\text{s}}}{0.018 \frac{\text{m}^3}{\text{s}} + 0.041 \frac{\text{m}^3}{\text{s}}} = 0.31 \quad (215)$$

The fraction of gas in the pipeline is therefore $1 - 0.31 = 0.69$ and the total gas volume in the pipeline is

$$V_G = A L (1 - \lambda_L) = 0.067 \text{ m}^2 \cdot 1000 \text{ m} \cdot 0.69 = 46.23 \text{ m}^3 \quad (216)$$

Assuming the ideal gas law $pV = nRT$, the rate of increase of pressure in the pipeline is

$$\left(\frac{dp}{dt} \right)_{\text{pipeline}} = \frac{dn}{dt} \frac{RT}{V_G} \quad (217)$$

The molar flowrate is

$$\frac{dn}{dt} = \frac{Q_G}{V_m} \quad (218)$$

where V_m is the gas molar volume which can be calculated using the ideal gas law

$$\frac{V}{n} = V_m = \frac{RT}{P} \quad (219)$$

The rate of increase of pressure is therefore

$$\left(\frac{dp}{dt} \right)_{\text{pipeline}} = \frac{P Q_G}{V_G} = \frac{20 \text{ bar} \cdot 0.041 \frac{\text{m}^3}{\text{s}}}{46.23 \text{ m}^3} = 0.018 \frac{\text{bar}}{\text{s}} \quad (220)$$

The rate of pressure increase as liquid fills the riser is larger than the rate of pressure increase in the gas section of the pipeline

$$0.02373 \frac{\text{bar}}{\text{s}} > 0.018 \frac{\text{bar}}{\text{s}} \quad (221)$$

Therefore one of the main criteria for severe slugging is satisfied.

[Question end]

Q.43**Question 43**

Given the following information for a horizontal multi-phase pipeline, calculate the expected pressure loss in bara using the Homogeneous Flow Model.

| | | |
|----------------|----------|----------------------------------|
| Liq density | 860 | kg m^{-3} |
| Gas density | 25 | kg m^{-3} |
| Liq viscosity | 0.0013 | $\text{kg m}^{-1} \text{s}^{-1}$ |
| Gas viscosity | 0.000012 | $\text{kg m}^{-1} \text{s}^{-1}$ |
| Mass flow rate | 50 | kg s^{-1} |
| Pipe i.d. | 0.2 | m |
| Pipe length | 1000 | m |
| Vapour quality | 0.03 | Mass fraction |

Solution:

First, calculate the pipe cross sectional area

$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot 0.2^2}{4} = 0.314 \text{ m}^2 \quad (222)$$

Using the category 1 Homogeneous Flow Model the two-phase pressure drop is calculated as

$$\Delta P = \frac{2 f_{2p} \rho_n v_m^2 L}{D} \quad (223)$$

where ρ_n is the no-slip two-phase density, v_m is the no-slip mixture velocity, and f_{2p} is the two-phase friction factor which requires the two-phase Reynolds number which is defined as

$$Re_n = \frac{\rho_n v_m D}{\mu_n} \quad (224)$$

where μ_n is the no-slip mixture viscosity. We, therefore, need to calculate v_m , ρ_n , and μ_n . Start by calculating the liquid and gas flowrates

$$Q_L = \frac{M(1-x)}{\rho_L} = \frac{50 \frac{\text{kg}}{\text{s}} (1-0.03)}{860 \frac{\text{kg}}{\text{m}^3}} = 0.056 \text{ m}^3 \text{ s}^{-1} \quad (225)$$

$$Q_G = \frac{Mx}{\rho_G} = \frac{50 \frac{\text{kg}}{\text{s}} \cdot 0.03}{25 \frac{\text{kg}}{\text{m}^3}} = 0.06 \text{ m}^3 \text{ s}^{-1} \quad (226)$$

Calculate the superficial velocities by dividing by the pipe cross sectional area

$$v_{SL} = \frac{Q_L}{A} = \frac{0.056 \frac{\text{m}^3}{\text{s}}}{0.314 \text{ m}^2} = 1.796 \text{ m s}^{-1} \quad (227)$$

$$v_{SG} = \frac{Q_G}{A} = \frac{0.06 \frac{\text{m}^3}{\text{s}}}{0.314 \text{ m}^2} = 1.911 \text{ m s}^{-1} \quad (228)$$

We can now calculate the mixture velocity

$$v_m = v_{SL} + v_{SG} = 1.796 + 1.911 = 3.707 \text{ m s}^{-1} \quad (229)$$

as well as the no-slip liquid hold-up

$$\lambda_L = \frac{v_{SL}}{v_m} = \frac{1.796 \frac{\text{m}}{\text{s}}}{3.707 \frac{\text{m}}{\text{s}}} = 0.4845 \quad (230)$$

This allows us to calculate the no-slip mixture density and viscosity

$$\rho_n = \lambda_L \rho_L + (1 - \lambda_L) \rho_G = 0.4845 \cdot 860 \frac{\text{kg}}{\text{m}^3} + (1 - 0.4845) \cdot 25 \frac{\text{kg}}{\text{m}^3} = 429.57 \text{ kg m}^{-3} \quad (231)$$

$$\mu_n = \lambda_L \mu_L + (1 - \lambda_L) \mu_G = 0.4845 \cdot 0.0013 \frac{\text{kg}}{\text{m s}} + (1 - 0.4845) \cdot 0.000012 \frac{\text{kg}}{\text{m s}} = 0.000636 \text{ kg m}^{-1} \text{ s}^{-1} \quad (232)$$

We now have the information available to calculate the Reynolds number

$$Re_n = \frac{\rho_n v_m D}{\mu_n} = \frac{429.57 \frac{\text{kg}}{\text{m}^3} \cdot 3.707 \frac{\text{m}}{\text{s}} \cdot 0.2 \text{ m}}{0.000636 \frac{\text{kg}}{\text{m s}}} = 500697 \quad (233)$$

We then calculate the friction factor using the Blasius equation

$$f_{2p} = 0.079 Re_n^{-0.25} = 0.079 \cdot 500697^{-0.25} = 0.00297 \quad (234)$$

The predicted pressure drop in the pipeline is therefore

$$\Delta P = \frac{2 f_{2p} \rho_n v_m^2 L}{D} = \frac{2 \cdot 0.00297 \cdot 429.57 \frac{\text{kg}}{\text{m}^3} \cdot (3.707 \frac{\text{m}}{\text{s}})^2 \cdot 1000 \text{ m}}{0.2 \text{ m}} = 1.753 \text{ bar} \quad (235)$$

[Question end]

Q.44 Question 44

7: Flow assurance

- a) Oil production from a vertical riser is enhanced by injecting gas at the riser base. A series of tests is conducted where the injected gas rate is varied. The oil production trend observed indicates an optimal gas rate which maximises oil production. If gas rates are reduced from the optimal gas rate, oil production drops, similarly if gas rates are increased above the optimal rate, oil production drops. Explain this observation and clearly state any assumptions you make. **[6 marks]**

Solution:

Oil flowrate will be a function of the pressure drop in the riser. For vertical flow there will be two components influencing the pressure drop-friction and elevation. Acceleration will be negligible².

As gas is first introduced, the oil flow rate increases as the gas lowers the density of the fluid in the column. Thus reducing the elevation pressure losses ($\rho g h$). Further increases in oil flow are seen as the gas rate is increased as the density is further reduced².

Although density is reducing which is reducing the elevation losses the frictional component of the pressure loss is increasing due to the increased system velocity. There comes a point where the benefits obtained from density reduction are overcome by friction. At this point further increases in gas rate result in an increased pressure drop and reduced oil rate. Hence the opposite trend is seen with respect to oil flowrate².

- b) Liquid and gas are flowing in a horizontal pipeline with inner diameter of 38.1 cm. The liquid rate is $0.0369 \text{ m}^3 \text{ s}^{-1}$. Using the Mandhane flow map in Fig. 12, calculate the gas rate in $\text{m}^3 \text{ s}^{-1}$ where there will be a transition from slug to annular flow. **[4 marks]**

Solution:

Calculating the pipeline cross-sectional area,

$$A = 3.141 \times 0.381^2 / 4 = 0.114 \text{ m}^2$$

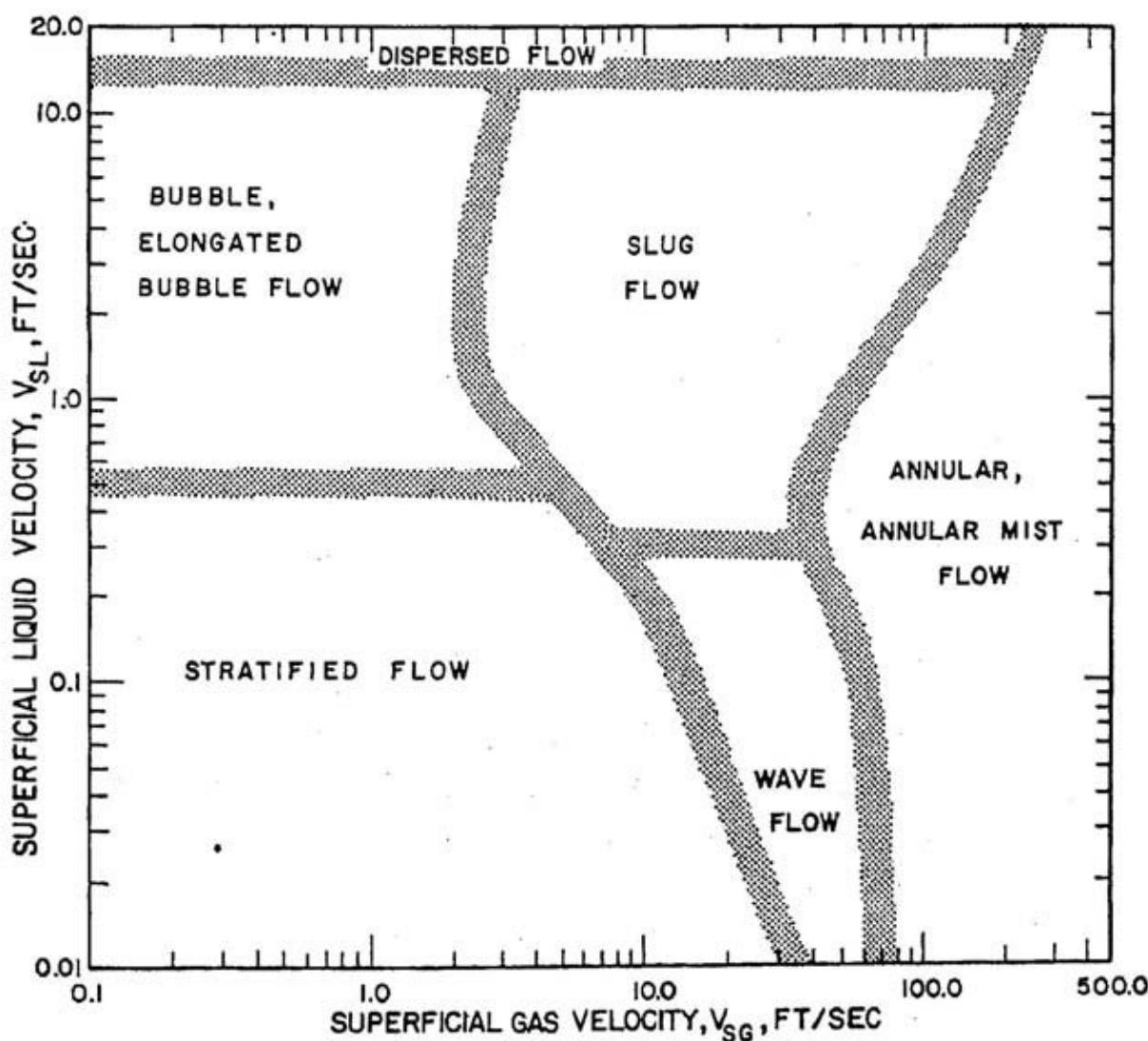


Figure 12: The Mandhane horizontal flow pattern map.

The liquid superficial velocity can then be calculated,

$$0.0369 / 0.114 = 0.324 \text{ m s}^{-1} = 1.06 \text{ ft s}^{-1}$$

[2/4] ✓ From the map we find that a transition occurs at a gas superficial velocity of approximately 50–70 ft s⁻¹, which is approximately 15.28–21.40 m s⁻¹. This is based on the thickness of the line; however, the size of the transition zone and any hysteresis it has are not well defined for such maps. The gas volumetric flow is then in the following range,

$$15.28 \times 0.114 = 1.74 \text{ m}^3 \text{ s}^{-1}$$

$$21.40 \times 0.114 = 2.44 \text{ m}^3 \text{ s}^{-1}$$

- [2/4] ✓
- c) What is a hydrate and what are the conditions required for their formation? [2 marks]

Solution:

A hydrate is an ice like substance where water reacts with small hydrocarbon molecules and water. Conditions promoting hydrate formation are low temperatures and high pressures.

[1/2]

[1/2]

- d) A pipeline is flowing with a light hydrocarbon gas and water. Given the following information, calculate the distance from the pipeline inlet to the point where hydrates will form. [8 marks]

- The hydrate formation temperature for the pipeline is 15 °C.
- Pipeline Overall Heat Transfer Coefficient (OHTC), $U = 5 \text{ W m}^{-2} \text{ }^{\circ}\text{C}^{-1}$ (based on pipe outer diameter).
- Pipeline inside diameter, $D_i = 0.2 \text{ m}$.
- Pipeline wall thickness, $W_t = 0.02 \text{ m}$.
- Gas/water mixture specific heat, $C_p = 2.8 \text{ kJ kg}^{-1} \text{ }^{\circ}\text{C}^{-1}$.
- Gas/water mixture mass flow rate, $\dot{m} = 10 \text{ kg s}^{-1}$.
- Ambient temperature, $T_a = 4^{\circ}\text{C}$.
- Pipeline inlet temperature, $T_i = 50^{\circ}\text{C}$.
- You should use a log-mean temperature difference, $\Delta T_{lm} = (\Delta T_A - \Delta T_B) / \ln\left(\frac{\Delta T_A}{\Delta T_B}\right)$ where A and B denote opposite ends of the exchanger/pipeline, when calculating the heat flux.

Solution:

The heat flux from the pipe is as follows:

$$Q = UA\Delta T_{lm} = \dot{m} C_p \Delta T_f$$

where ΔT_f is the fluid temperature change. The pipe area is a function of length:

$$A = 3.141 D_o L = 0.754 L \text{ m}^2$$

[2/8] ✓ where $D_o = 0.2 + 2 \times 0.02 = 0.24 \text{ m}$. Assuming the fluid reaches 15 °C at one end, the LMTD is,

$$\Delta T_{lm} = ((50 - 4) - (15 - 4)) / \ln\left(\frac{50 - 4}{15 - 4}\right) = 35 / \ln(4.182) = 24.5 \text{ }^{\circ}\text{C}$$

[2/8] ✓ The sensible heat change is,

$$\dot{m} C_p \Delta T_f = 10 \times 2800 \times (50 - 15) = 980000 \text{ J s}^{-1}$$

[2/8] ✓ Equating this with the heat flux to calculate the area

$$A = 980000 / (5 \times 24.5) = 8000 \text{ m}^2$$

[2/8] Thus $L = A / 0.754 = 8000 / 0.754 = 10610 \text{ m}$.

[Question total: 20 marks]

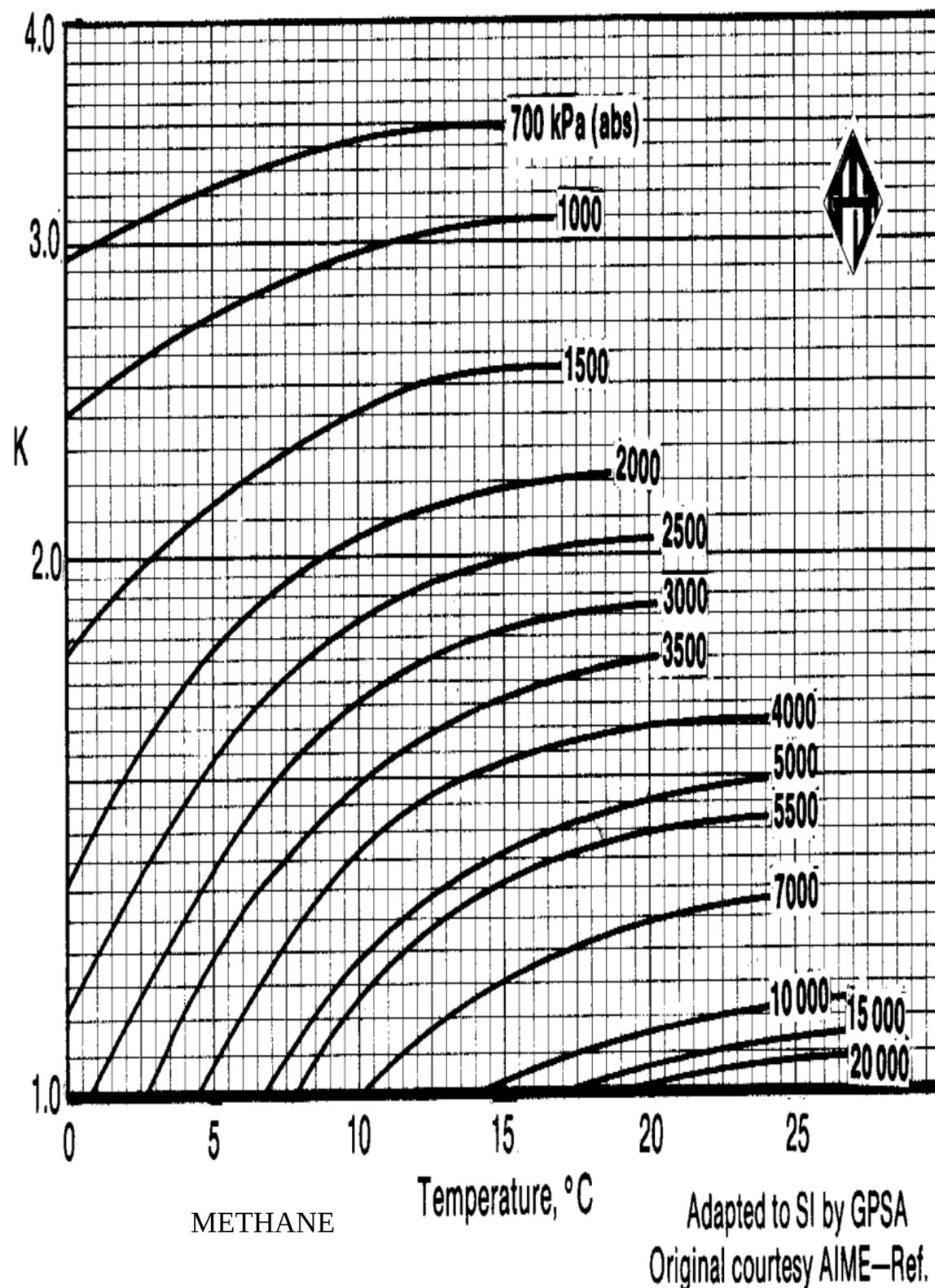
8 8: Hydrates, wax, and erosion

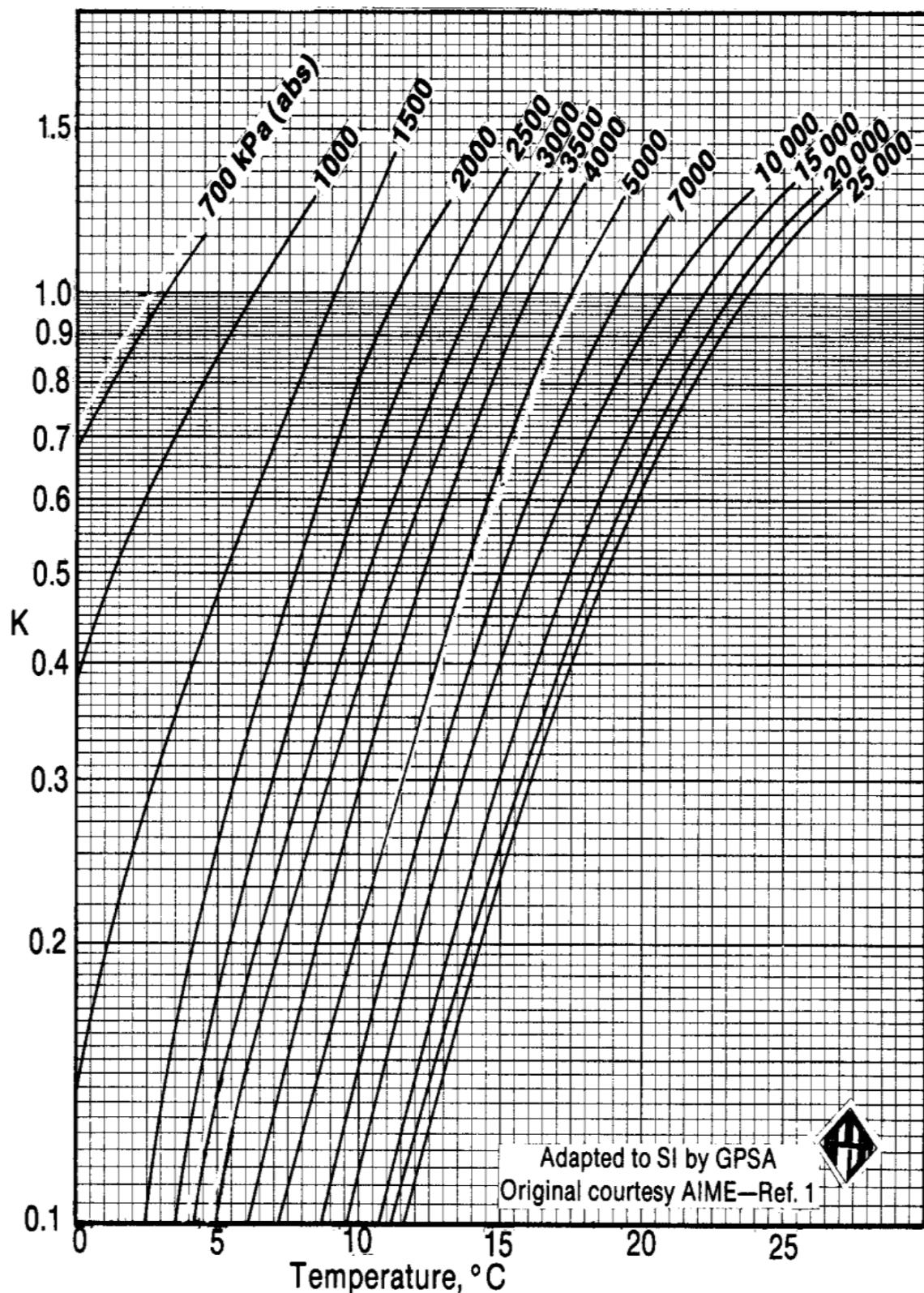
Q.45 Question 45

The following mixture is flowing in a subsea pipeline.

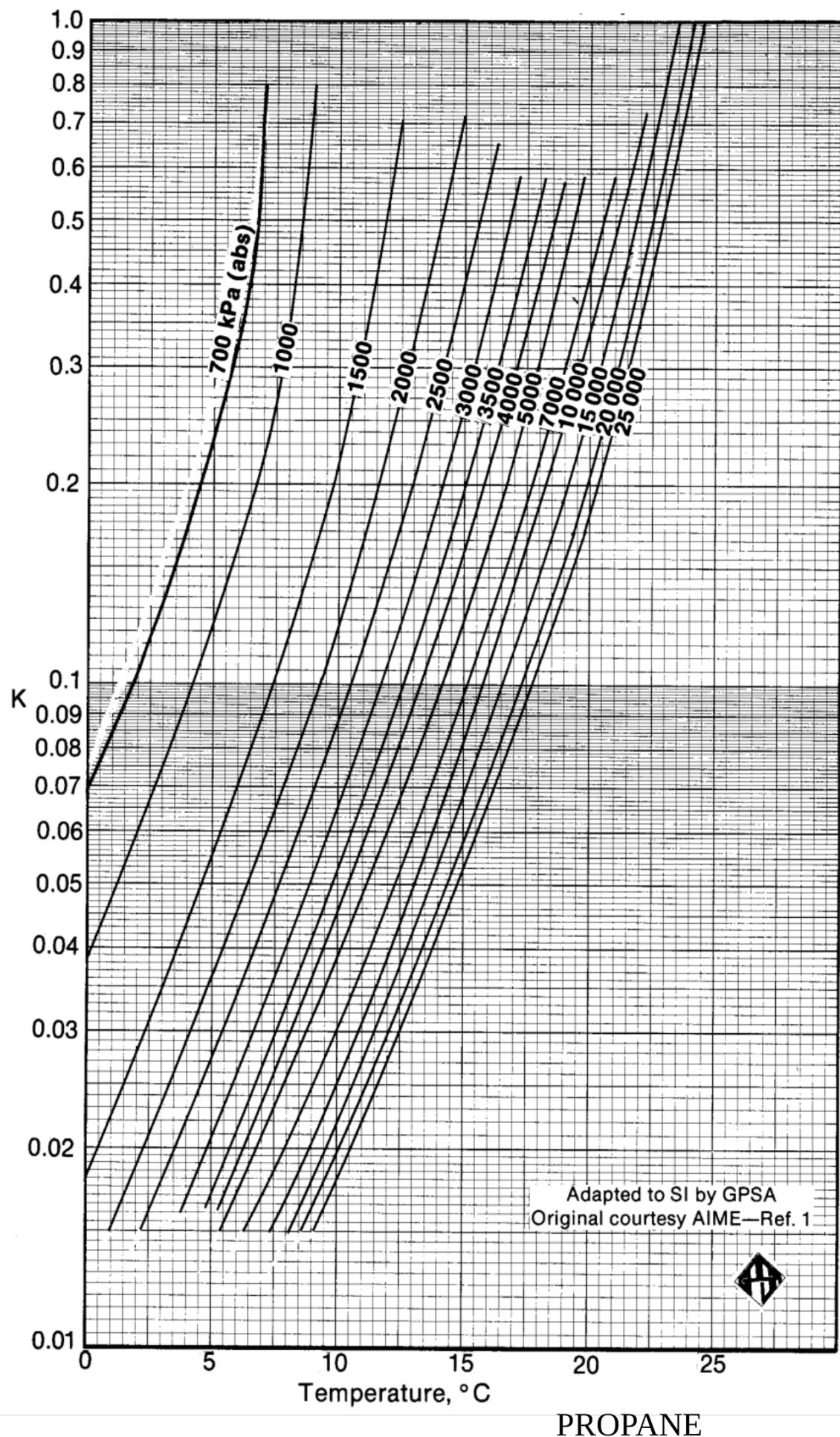
| Component | Mol fraction |
|-----------|--------------|
| Methane | 0.6 |
| Ethane | 0.3 |
| Propane | 0.1 |

Calculate the hydrate formation temperature for a pipeline pressure of 40 bara. Use the K value figures and compare with the GPSA charts.

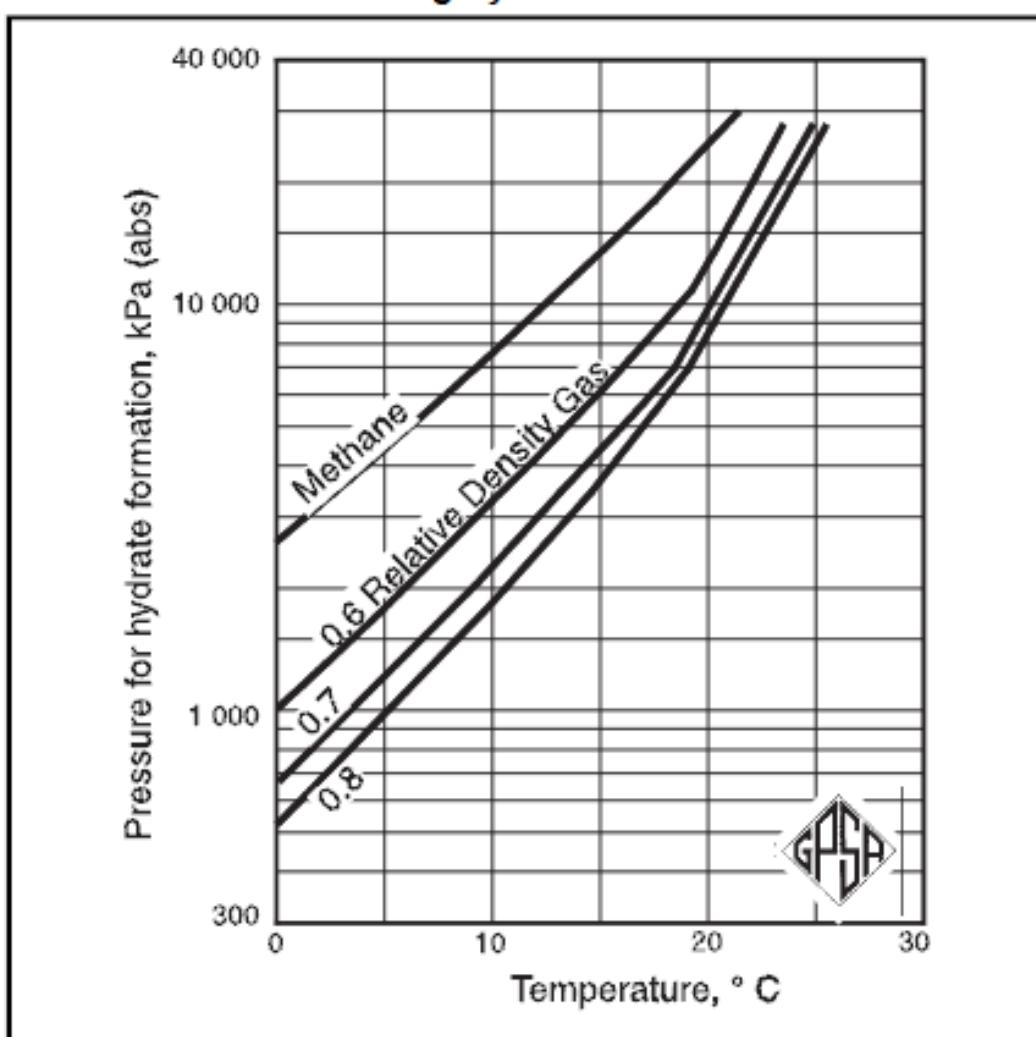




ETHANE



Pressure-Temperature Curves for Predicting Hydrate Formation


Solution:
Using the GPSA chart

The molecular weights of the components area

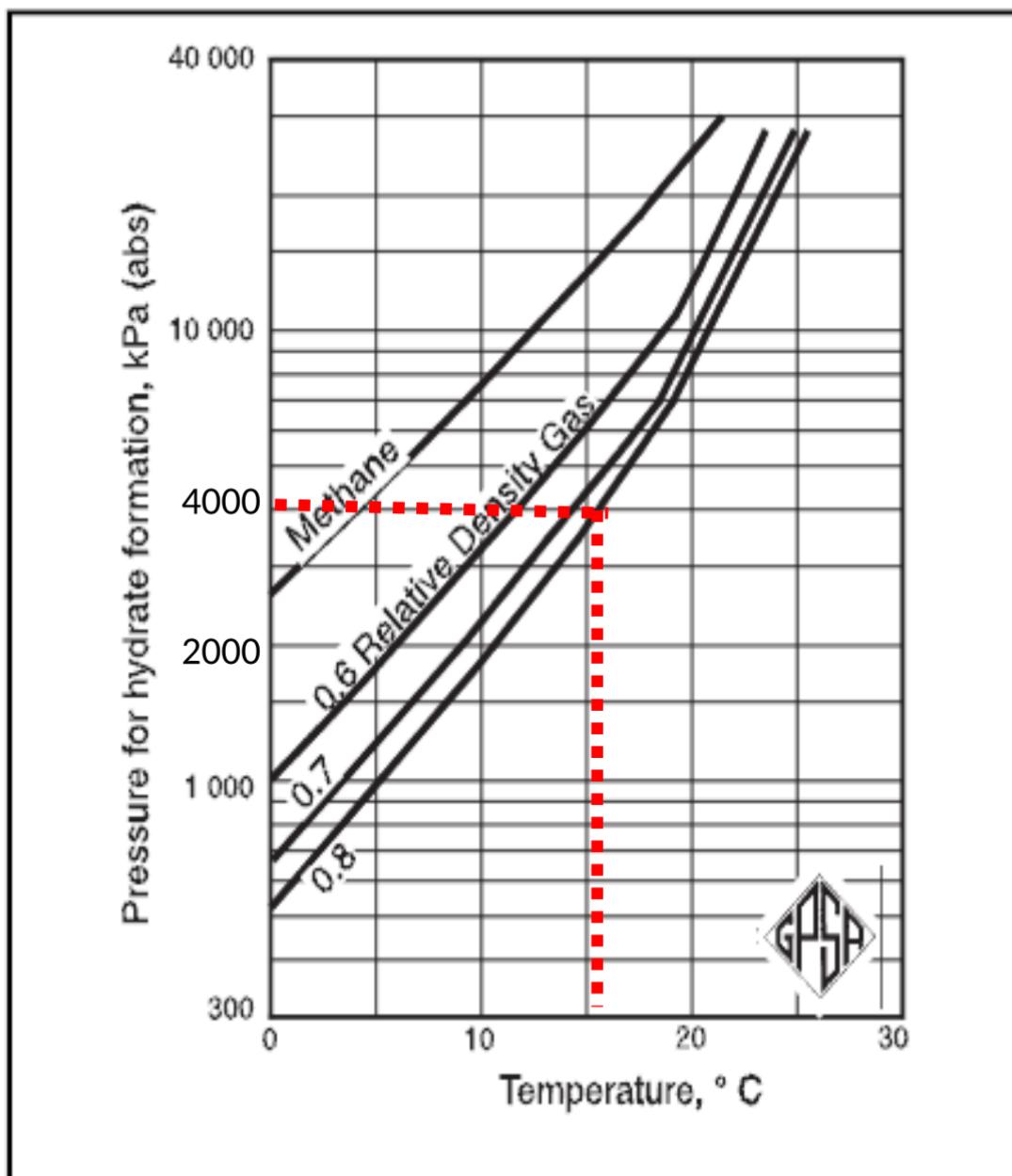
- Methane: 16 g mol^{-1}
- Ethane: 30 g mol^{-1}
- Propane: 44 g mol^{-1}

Calculate the average molecular weight of the mixture

$$M_{\text{ave}} = 0.6 \cdot 16 \frac{\text{g}}{\text{mol}} + 0.3 \cdot 30 \frac{\text{g}}{\text{mol}} + 0.1 \cdot 44 \frac{\text{g}}{\text{mol}} = 23 \frac{\text{g}}{\text{mol}} \quad (236)$$

Calculate the relative gas density = $M_{\text{ave}}/M_{\text{air}} = 23/29 = 0.793$. Obtain the hydrate formation temperature using the GPSA chart with a relative gas density = 0.793 and a pressure of 4000 kPa.

Pressure-Temperature Curves for Predicting Hydrate Formation



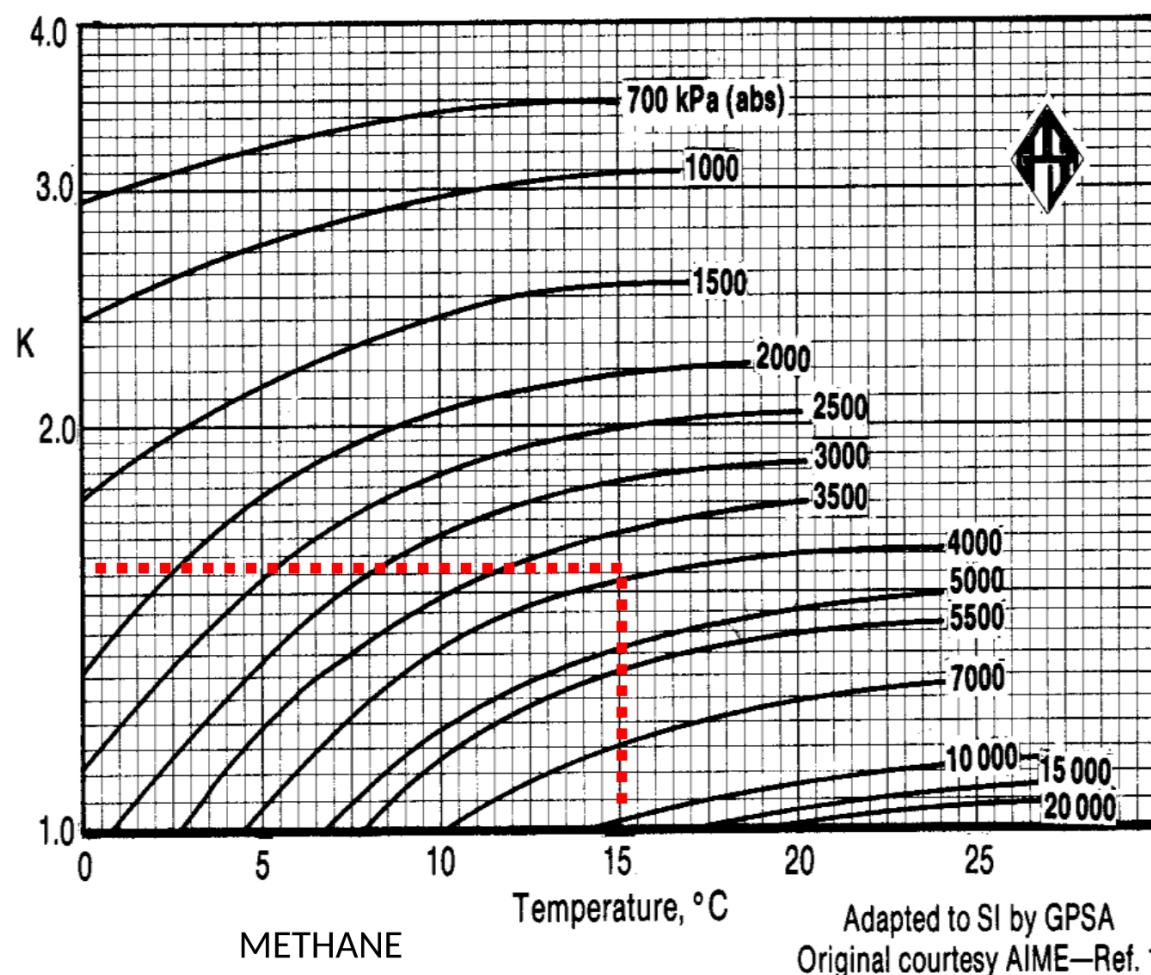
The chart gives a hydrate formation temperature of 16 °C.

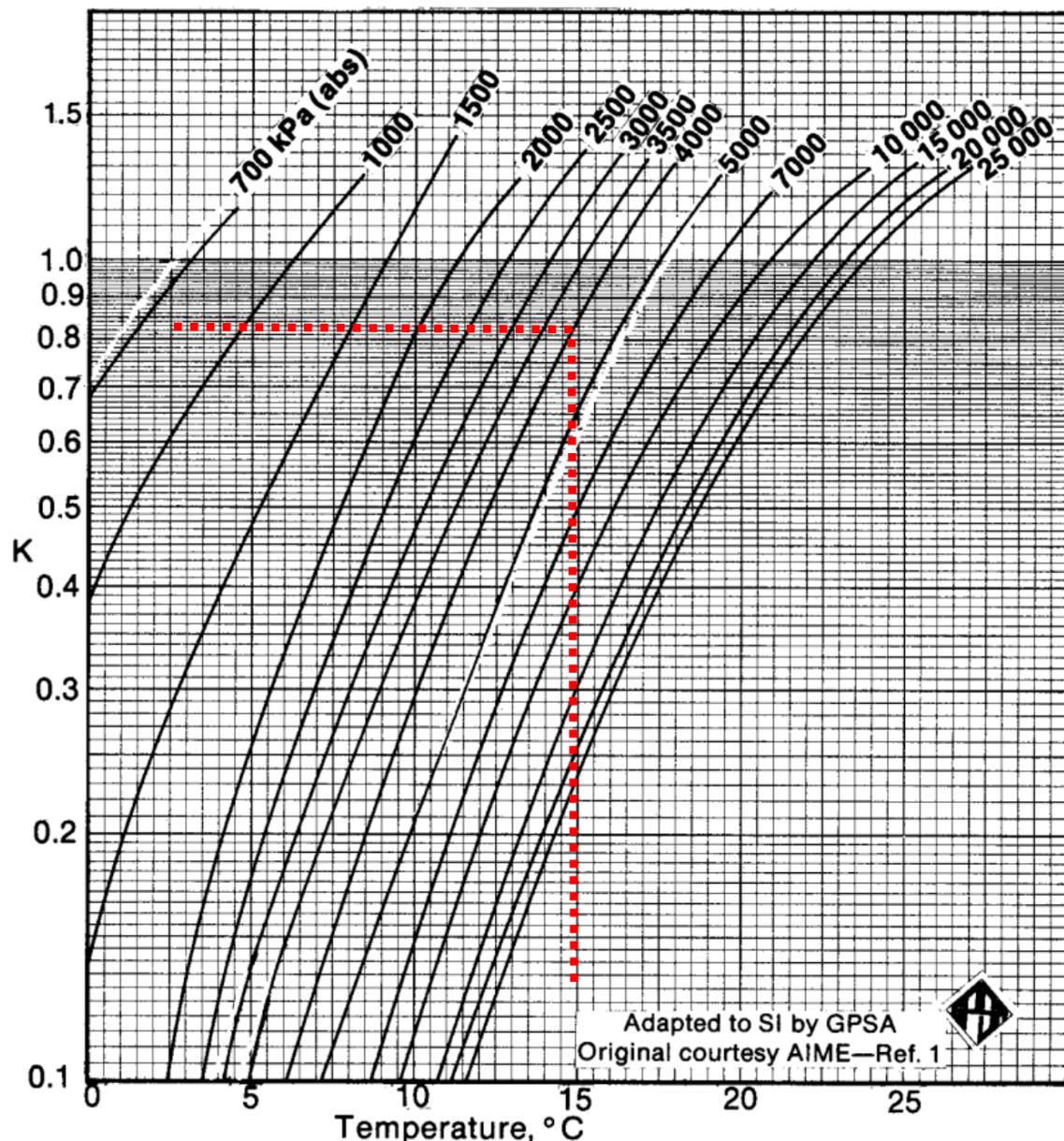
Using the K-value chart

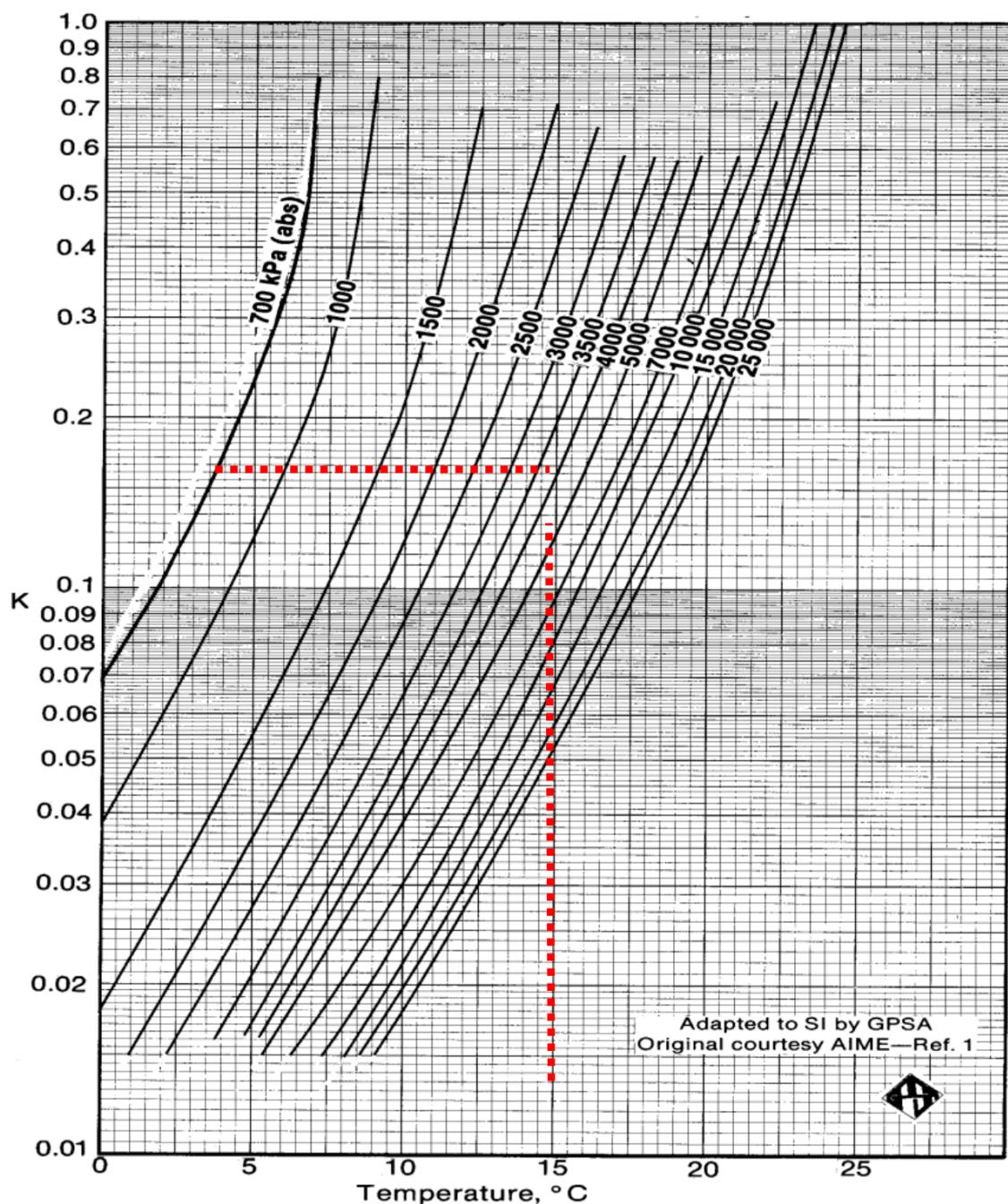
At the hydrate formation the mixture will satisfy

$$\sum_{i=1}^n \frac{y_i}{K_i} = 1 \quad (237)$$

Note the similarity to dew point calculation. Start with an initial guess for the hydrate formation temperature e.g. $T = 15$ °C. Obtain K values for each component using a pressure of 4000 kPa.







These K-values will result in

| 15 °C | K_i | y_i | y_i/K_i |
|---------|-------|-------|-----------|
| Methane | 1.55 | 0.6 | 0.387097 |
| Ethane | 0.85 | 0.3 | 0.352941 |
| Propane | 0.175 | 0.1 | 0.571429 |
| | | Sum | 1.311467 |

In this case, $\sum_{i=1}^n \frac{y_i}{K_i} > 1$. Repeat this exercise for a higher temperature (in this case a higher temperature will result in a sum which is closer to 1). If we choose $T = 17$ °C we get

| 17 °C | K_i | y_i | y_i/K_i |
|---------|-------|-------|-----------|
| Methane | 1.57 | 0.6 | 0.382166 |
| Ethane | 1.15 | 0.3 | 0.26087 |
| Propane | 0.25 | 0.1 | 0.4 |
| | | Sum | 1.043035 |

Here, $\sum_{i=1}^n \frac{y_i}{K_i} = 1.04$. Therefore the hydrate formation temperature will be slightly larger than $T = 17$ °C. This is slightly greater than the temperature predicted using the GPSA chart.

[Question end]

Q.46

Question 46

The hydrate formation temperature of a hydrocarbon and water mixture flowing in a subsea pipeline has been measured at 18 °C. The pipeline conditions are:

- Pipeline length $L = 15$ km
- Pipe inner diameter $D = 0.292$ m
- Pipe wall thickness = 0.0127 m
- Mass flowrate $m = 15$ kg s⁻¹
- Inlet temperature $T_1 = 70$ °C
- Sea temperature $T_{\text{out}} = 4$ °C
- Fluid specific heat $C_p = 1745$ J kg⁻¹ K⁻¹

What is the minimum level of insulation (U value) that would be required to ensure that the pipeline contents will not enter hydrate forming conditions?

Solution:

The overall heat transfer coefficient can be determined using a heat balance

$$mC_p(T_1 - T_2) = UA\Delta T_{lm} \quad (238)$$

where ΔT_{lm} is the log-mean temperature difference (in this case the sea temperature outside the pipe is constant)

$$\Delta T_{lm} = \frac{(T_1 - T_{\text{out}}) - (T_2 - T_{\text{out}})}{\ln \left(\frac{T_1 - T_{\text{out}}}{T_2 - T_{\text{out}}} \right)} \quad (239)$$

Substituting the log-mean temperature difference in to the heat balance equation gives the following expression for U

$$U = \frac{mC_p}{A} \ln \left(\frac{T_1 - T_{\text{out}}}{T_2 - T_{\text{out}}} \right) \quad (240)$$

Calculate U using $T_2 = 18$ °C, i.e. the pipeline contents will remain above the hydrate formation temperature of 18 °C

$$U = \frac{15 \frac{\text{kg}}{\text{s}} \cdot 1745 \frac{\text{J}}{\text{kg K}}}{\pi \cdot (0.292 \text{ m} + 2 \cdot 0.0127 \text{ m}) \cdot 15000 \text{ m}} \ln \left(\frac{70 \text{ °C} - 4 \text{ °C}}{18 \text{ °C} - 4 \text{ °C}} \right) \quad (241)$$

$$= 2.71 \frac{\text{W}}{\text{m}^2 \text{ K}} \quad (242)$$

[Question end]

Q.47

Question 47

Calculate the time taken for a subsea pipeline to cool to hydrate forming conditions after flow is stopped. Consider two sections of the pipe, one gas filled and the other mainly liquid filled. Each section of the pipe is 10 m long. The pipe ID and OD are 0.17 m and 0.2 m respectively. Assume the pipe OHTC is $1.5 \text{ W m}^{-2} \text{ K}^{-1}$. The specific heat C_p of the gas and the liquid is $3100 \text{ J kg}^{-1} \text{ K}^{-1}$. The gas density is 90 kg m^{-3} and the liquid density is 840 kg m^{-3} . The initial temperature is 70°C and the surrounding sea temperature is 4°C . The hydrate forming temperature is 15°C .

Solution:

Rearrange the temperature decay equation

$$T = T_{\text{out}} + (T_{\text{initial}} - T_{\text{out}}) e^{\frac{UA}{mC_p} t} \quad (243)$$

to obtain an expression for the time

$$t = -\frac{mC_p}{UA} \ln \left(\frac{T - T_{\text{out}}}{T_{\text{initial}} - T_{\text{out}}} \right) \quad (244)$$

liquid-filled section

The total mass of liquid in the section is

$$m_L = \rho_L V = 840 \frac{\text{kg}}{\text{m}^3} \cdot \frac{\pi \cdot (0.17 \text{ m})^2 \cdot 10 \text{ m}}{4} \quad (245)$$

$$= 190.66 \text{ kg} \quad (246)$$

Calculate the time t using a final temperature $T = 15^\circ\text{C}$, i.e. the hydrate formation temperature

$$t = -\frac{190.66 \text{ kg} \cdot 3100 \frac{\text{J}}{\text{kg K}}}{1.5 \frac{\text{W}}{\text{m}^2 \text{K}} \cdot (\pi \cdot 0.2 \text{ m} \cdot 10 \text{ m})} \ln \left(\frac{15^\circ\text{C} - 4^\circ\text{C}}{70^\circ\text{C} - 4^\circ\text{C}} \right) \quad (247)$$

$$= 112365 \text{ s} = 31.2 \text{ hrs} \quad (248)$$

gas-filled section

The total mass of gas in the section is

$$m_G = \rho_G V = 90 \frac{\text{kg}}{\text{m}^3} \cdot \frac{\pi \cdot (0.17 \text{ m})^2 \cdot 10 \text{ m}}{4} \quad (249)$$

$$= 20.43 \text{ kg} \quad (250)$$

Again, calculate the time t using a final temperature $T = 15^\circ\text{C}$, i.e. the hydrate formation temperature

$$t = -\frac{20.43 \text{ kg} \cdot 3100 \frac{\text{J}}{\text{kg K}}}{1.5 \frac{\text{W}}{\text{m}^2 \text{K}} \cdot (\pi \cdot 0.2 \text{ m} \cdot 10 \text{ m})} \ln \left(\frac{15^\circ\text{C} - 4^\circ\text{C}}{70^\circ\text{C} - 4^\circ\text{C}} \right) \quad (251)$$

$$= 12040 \text{ s} = 3.3 \text{ hrs} \quad (252)$$

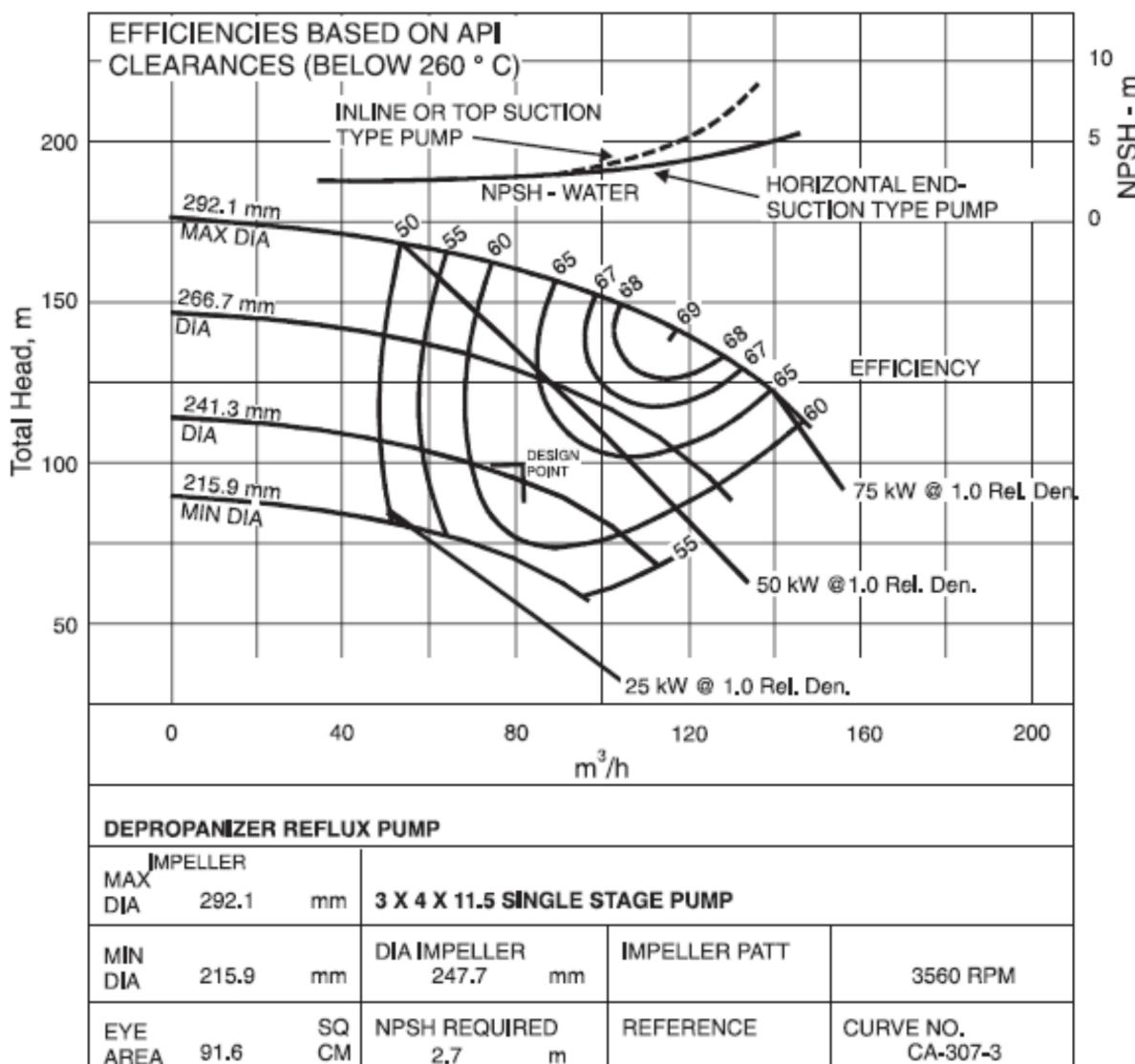
[Question end]

9 9: Energy efficiency

Q.48

Question 48

Verify the affinity laws using the following pump characteristic. The duty is water and the pump has a 215.9 mm impeller. The flow rate is $80 \text{ m}^3 \text{ hr}^{-1}$. The impeller is changed to 266.7 mm. Calculate the new head using the affinity laws and check against values in the figure. Similarly for power.



Solution:

The affinity laws are

$$\frac{q_1}{q_2} = \left(\frac{N_1}{N_2} \right) \cdot \left(\frac{d_1}{d_2} \right)^2 \quad (253)$$

$$\frac{H_1}{H_2} = \left(\frac{N_1}{N_2} \right)^2 \cdot \left(\frac{d_1}{d_2} \right)^2 \quad (254)$$

$$\frac{W_1}{W_2} = \left(\frac{N_1}{N_2} \right)^3 \cdot \left(\frac{d_1}{d_2} \right)^3 \quad (255)$$

where q is the flow rate, N is the rotational speed, d is the impeller diameter, H is the pump head, and W is the pump head.

From the chart, at a flow rate of $80 \text{ m}^3 \text{ hr}^{-1}$ and impeller diameter of 215.9 mm, the pump power and head values are 27 kW and 73 m respectively.

From the chart, the new pump power and head values at an impeller size of 266.7 mm are 47 kW and 128 m respectively.

Using the affinity laws, the power and head values area

$$H_1 = 73 \text{ m} \cdot \left(\frac{266.7 \text{ mm}}{215.9 \text{ mm}} \right)^2 = 111.4 \text{ m} \quad (256)$$

$$W_1 = 27 \text{ kW} \cdot \left(\frac{266.7 \text{ mm}}{215.9 \text{ mm}} \right)^3 = 50.9 \text{ kW} \quad (257)$$

[Question end]

Q.49

Question 49

Liquid propane ($\gamma = 0.502$) at its bubble point is pumped from a pressure vessel (operating pressure = 10 bara) by a centrifugal pump. The elevation of the liquid level in the suction vessel is 5 m above datum. The elevation of the pump suction nozzle is 1 m above datum. The friction loss in the suction line is 0.075 bar.

- What is the available NPSH?
- The pump has a NPSH requirement of 5 m. How could you change the equipment configuration to achieve a NPSH of 5 m? Due to other restrictions the vessel cannot be elevated.

Solution:

- *What is the available NPSH?*

The avaiable NPSH is calculated using

$$NPSH_{\text{available}} = \frac{10.19 (p_s - p_v - \Delta p_f)}{\gamma} + H_s - H_m - \frac{v^2}{2g} - H_{ac} \quad (258)$$

where p_s is the pressure inside the suction vessel, p_v is the vapour pressure of the liquid at pump suction, Δp_f is the friction loss in the suction line, γ is the liquid specific gravity, H_s

is the height difference between the drawdown line in the suction vessel and the centerline of pump suction, H_m is the allowance for the extra available head above minimum, v is the liquid velocity at pump entry, and H_{ac} is the acceleration head (relevant for reciprocating pumps).

In this case

- The velocity term is assumed to be negligible, $v = 0 \frac{m}{s}$
- H_{ac} is only valid for reciprocating pumps ($H_{ac} = 0$ m)
- There is no allowance for extra available head ($H_m = 0$ m)
- At the bubble point $p_s = p_v$

Calculate the NPSH

$$NPSH_{\text{available}} = \frac{10.19 (-\Delta p_f)}{\gamma} + H_s \quad (259)$$

$$= \frac{10.19 \cdot -0.075 \text{ bar}}{0.502} + (5 \text{ m} - 1 \text{ m}) \quad (260)$$

$$= 2.48 \text{ m} \quad (261)$$

- *The pump has a NPSH requirement of 5 m. How could you change the equipment configuration to achieve a NPSH of 5 m? Due to other restrictions the vessel cannot be elevated.*

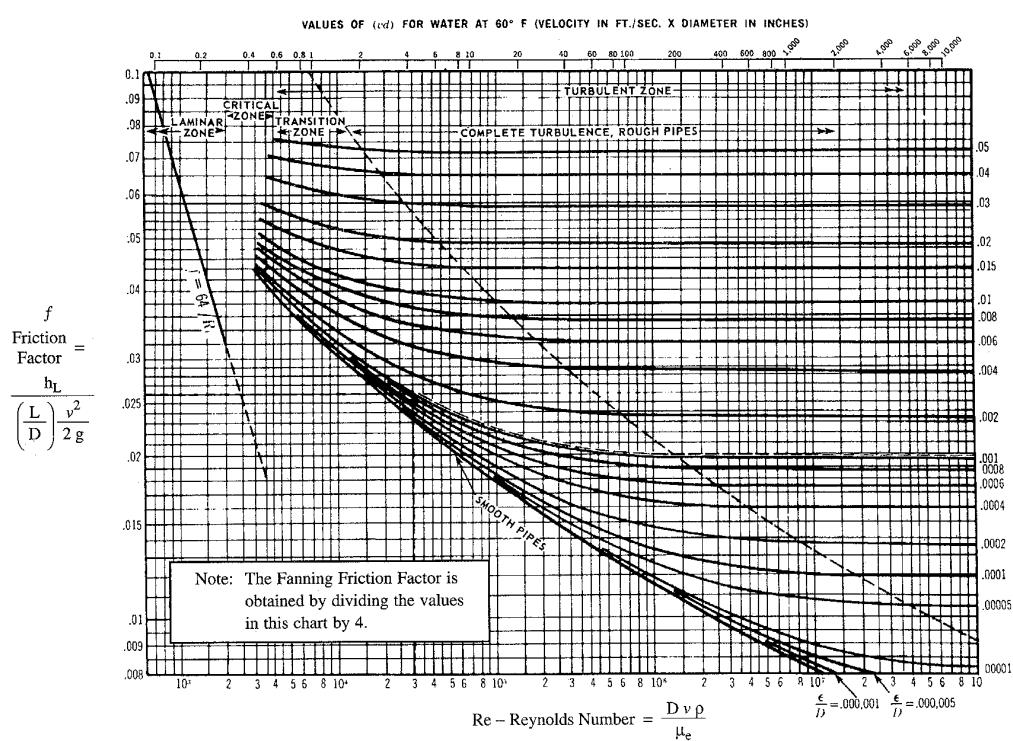
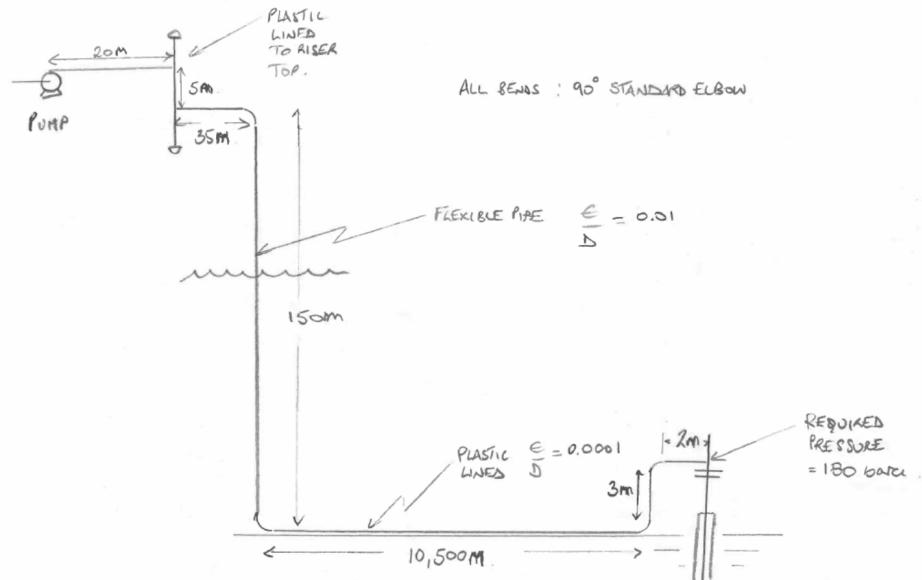
Adding a cooler in the pump suction line will reduce the vapour pressure while the suction pressure remains constant resulting in a greater available NPSH.

[Question end]

Q.50

Question 50

A plastic lined subsea water injection flowline is required to deliver 80,000 barrels per day of treated sea water to a subsea manifold at a pressure of 180 bara and temperature of 10 °C. The water pipeline route is as shown. The topsides piping is also plastic lined but the riser is flexible pipe. The pipe ID is 10.75 inch throughout. Viscosity and density of sea water at 10 °C is 1.31 cP and 1020 kg m⁻³ respectively. Calculate the required water injection pump discharge pressure and pump power requirements if the pump suction pressure is 10 bara. The pump efficiency is 75 %.



| Description | | | | Equivalent Length in Pipe Diameters (L/D) |
|--|---|--|---|--|
| Globe Valves | Stem perpendicular to run | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 340 450 |
| | Y-pattern | Stem 60° from run of pipe Stem 45° from run of pipe | 100% Open 100% Open | 175 145 |
| Angle Valves | | No Obstruction Wing or Pin guided disc | 100% Open 100% Open | 145 200 |
| Gate Valves | Wedge Disc Double Disc, Plug Disc | | 100% Open 75% Open 50% Open 25% Open | 13 35 160 900 |
| | Pulp Stock | | 100% Open 75% Open 50% Open 25% Open | 17 50 260 1200 |
| Conduit Pipe Line Gate, Ball and Plug Valves | | | 100% Open | 3 |
| Check Valves | Conventional Swing Clearway Swing Globe Lift or Stop; Stem Perpendicular to Run or Y-pattern Angle lift or Stop In-Line Ball | | | 135 50 same as globe same as globe 150 |
| Foot Valves with Strainer | | Poppet lift-type disc Leather-hinger disc | | 420 75 |
| Butterfly Valves (8-inch and larger) | | | | 40 |
| Cocks | Straight-through | Rectangular plug port area equal to 100% of pipe area | | 18 |
| | Three-way | Rectangular plug port area equal to 80% of pipe area | Straight through Through branch | 44 140 |
| Fittings | 90° Standard Elbow 45° Standard Elbow 90° Long Radius Elbow | | | 30 16 20 |
| | 90° Street Elbow 45° Street Elbow Square Corner Elbow | | | 50 26 57 |
| | Standard Tee | Flow through run Flow through branch | | 20 60 |
| | Close Pattern Return Bend | | | 50 |

[Question end]

Q.51 Question 51

A waste water treatment plant is considering a new pumping station. The plant owner has requested quotes from the approved pump supplier for supplying pumps with differing efficiencies. The costs received are:

| Efficiency % | Pump station cost £ |
|--------------|---------------------|
| 65 | 3,400,000 |
| 70 | 5,400,000 |
| 75 | 6,310,000 |
| 80 | 7,200,000 |
| 85 | 8,875,000 |

The pumps are electric motor driven. At 65 % efficiency the pump power requirement will be 10 MW. The plant design life is 15 years, the required availability is 0.95 and the cost of electricity is $0.025 \text{ £ kW}^{-1} \text{ hr}^{-1}$. What is the optimal efficiency to minimise life of field costs? Assume a discount rate of 3 %.

[Question end]

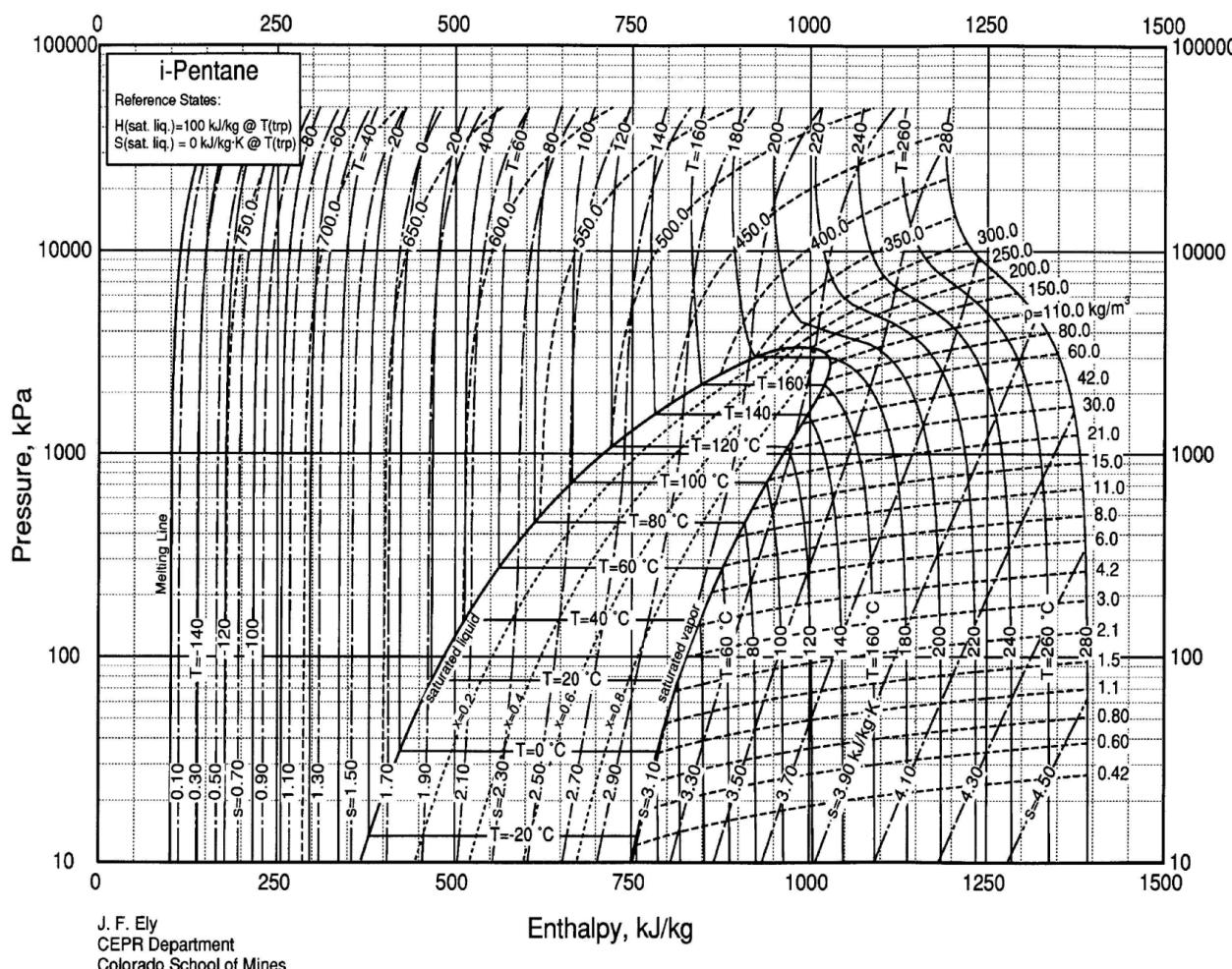
Q.52**Question 52**

A high pressure gas condensate well flows at $30,000 \text{ kg hr}^{-1}$. The arrival pressure and temperature are 650 bara and 160°C . The pressure is dropped across a choke valve to 100 bara. If the choke were replaced by an expander how much energy could be produced? The inlet and outlet compressibilities are 0.85 and 0.9 respectively. The gas k factor is 1.18. Assume a polytropic efficiency of 70 %. The gas molecular weight is 28.

[Question end]

Q.53**Question 53**

i-pentane is to be utilised in an Organic Rankine Cycle (ORC). 100,000 BPD of hot water is available at 100°C . Cooling water is available at 20°C . Use the attached Mollier chart and estimate the potential for power recovery in kW.



[Question end]

Q.54**Question 54**

15 kg s^{-1} of methane is to be compressed from 5 bara to 15 bara. The suction temperature to the compressor is 30°C .

- Compare a single stage compressor power requirements with a two stage compressor with intercooling to 30°C . Assume the compressor polytropic is 77 % and equal compression ratios across the two stage option.

- The single stage compressor is in operation. Conditions are steady with a constant compression ratio, but the following outlet temperatures are recorded. The inlet temperature remains at 30 °C. Quantify the corresponding compressor polytropic efficiencies. What could cause this trend?

| Year | Outlet temperature °C |
|------|-----------------------|
| 1 | 150 |
| 2 | 161 |
| 3 | 166 |

[Question end]

Q.55 Question 55

A reciprocating compressor has 6 cylinders each with a total cylinder volume of 0.002 m³. The cylinder diameter is 8 cm. The cylinder has a full stroke of 18 cm but the design has left a clearance distance of 4 mm. Quantify how this will affect the volumetric efficiency.

[Question end]

Q.56 Question 56

An oilfield with a 20 year life has the following energy demand

| Year | Load MW | Year | Load MW |
|------|---------|------|---------|
| 1 | 22 | 11 | 24 |
| 2 | 22 | 12 | 24 |
| 3 | 22 | 13 | 24 |
| 4 | 24 | 14 | 24 |
| 5 | 24 | 15 | 24 |
| 6 | 24 | 16 | 24 |
| 7 | 24 | 17 | 24 |
| 8 | 26 | 18 | 24 |
| 9 | 28 | 19 | 24 |
| 10 | 28 | 20 | 24 |

The choice of gas turbine to provide this power is between Solar Titan and a Solar Mars. Data for these two machines is shown.

| Platform Load MW | Solar Titan | | | |
|---------------------|-------------|--------------------------|------------------------|-------------------------------|
| | No Req'd | Load per Machine (MW) | Load (% of maximum) | Efficiency per Machine (%) |
| 10 | 1 | 10.0 | 80.3 | 29.8 |
| 12 | 1 | 12.0 | 96.4 | 31.1 |
| 14 | 2 | 7.0 | 56.2 | 25.3 |
| 16 | 2 | 8.0 | 64.2 | 27.1 |
| 18 | 2 | 9.0 | 72.3 | 28.6 |
| 20 | 2 | 10.0 | 80.3 | 29.8 |
| 22 | 2 | 11.0 | 88.3 | 30.6 |
| 24 | 2 | 12.0 | 96.4 | 31.1 |
| 26 | 3 | 8.7 | 69.6 | 28.2 |
| 28 | 3 | 9.3 | 74.9 | 29.1 |
| 30 | 3 | 10.0 | 80.3 | 29.8 |

| Platform Load MW | Solar Mars | | | |
|---------------------|------------|--------------------------|------------------------|-------------------------------|
| | No Req'd | Load per Machine (MW) | Load (% of maximum) | Efficiency per Machine (%) |
| 10 | 2 | 5.0 | 52.6 | 24.4 |
| 12 | 2 | 6.0 | 63.2 | 26.9 |
| 14 | 2 | 7.0 | 73.7 | 28.9 |
| 16 | 2 | 8.0 | 84.2 | 30.2 |
| 18 | 2 | 9.0 | 94.7 | 31.0 |
| 20 | 3 | 6.7 | 70.2 | 28.3 |
| 22 | 3 | 7.3 | 77.2 | 29.4 |
| 24 | 3 | 8.0 | 84.2 | 30.2 |
| 26 | 3 | 8.7 | 91.2 | 30.8 |
| 28 | 3 | 9.3 | 98.2 | 31.2 |
| 30 | 4 | 7.5 | 78.9 | 29.6 |

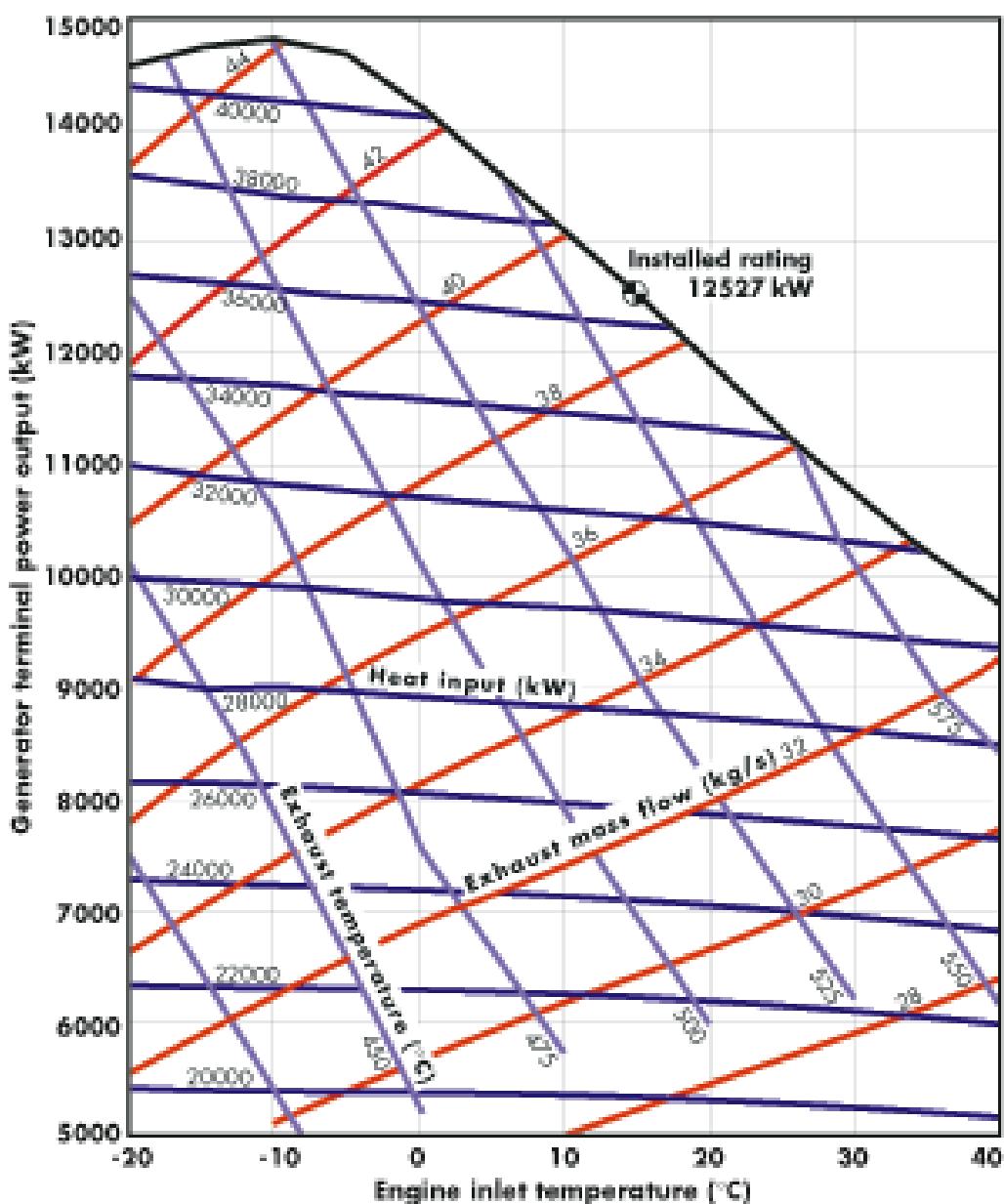
By conducting an annual review of machine efficiencies, determine which GT would be the most efficient over the field life.

[Question end]

Q.57 Question 57

A gas turbine iso rating is as shown.

- What full load power output would be achieved at an air temperature of 15 °C?
- What output could be achieved if the air temperature increased to 30 °C?
- What is the machine efficiency at 15 °C and 30 °C?



[Question end]

Q.58 Question 58

9: Energy efficiency

The following oil, gas, water separation and associated compression arrangement has been proposed for an offshore development. It is a single process train comprising two stages of three phase separation with interstage heating.

An export pump and cooler deliver the oil to an export pipeline. The export pump is driven by a fixed speed electric motor in a $2 \times 100\%$ parallel configuration. The pump efficiency at design conditions is 70%.

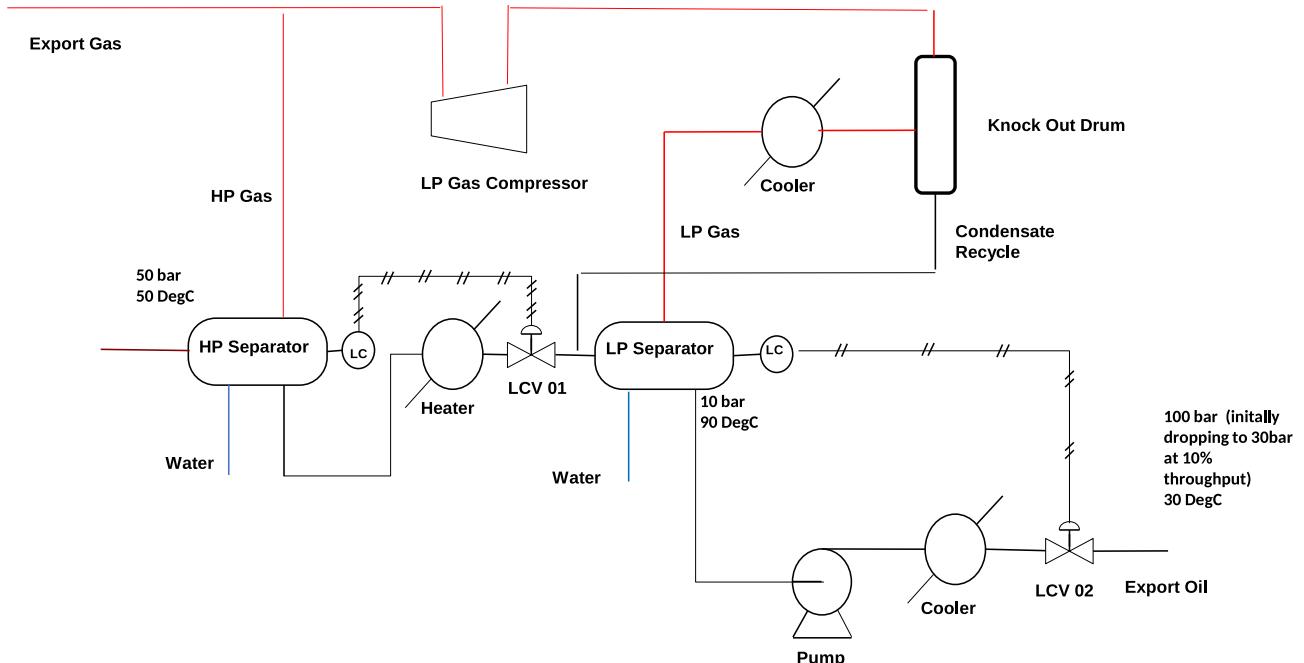
Large quantities of hot water are produced from each separator.

The interstage heater is an electric unit. Electrical supply to the heater, pump and compressor is from a simple cycle, gas turbine coupled to an alternator. The gas turbine is fuelled by process gas.

The system is designed for a 100% throughput but as the development ages, flowrates will drop progressively to 10% of design. The initial discharge requirement from the pump is 100 bara, but as the throughput reduces, the back pressure from the pipeline reduces requiring

less pump discharge pressure.

Gas from the low-pressure separator is cooled and passed to a knock out drum. The liquids from the drum are recycled to the inlet of the low pressure Separator. The gas is compressed in a $1 \times 100\%$ low pressure centrifugal compressor to a pressure equal to the high pressure Gas. The compressor is driven by a fixed speed electric motor and controlled by an anti-surge recycle system. The compressor polytropic efficiency is 70%.



Review this arrangement from an energy efficiency standpoint, identifying 10 topics which you would recommend require investigation before the design is finalised. For each topic identified give a short description of why you think it is worthy of consideration. **[20 marks]**

Solution:

2 mark for each for any of the 10 identified.

Export pump arrangement. Existing design will be inefficient at turndown conditions as the pumps will be far from their design duty. Consider other arrangements such as $4 \times 33\%$, $5 \times 25\%$. As above for compressors.

Variable speed drives for export pumps. Variable speed drives will minimise the requirement for recycle control which is inefficient. It will also allow the pumps to operate closer to their best efficiency point.

As above for compressor.

Pump efficiency appears low. Improved efficiency will reduce fuel gas usage.

As above for compressors.

Impeller change out for pumps to match turndown conditions. This would improve efficiency at turndown conditions.

As above for compressors.

Combined cycle rather than simple cycle gas turbine. This would result in a significant improvement in generation efficiency and reduction in the requirement for fuel gas requirement.

Combined heat and power. Use the heat from the gas turbine exhaust gas as process heating thus reducing electrical load.

Use the hot oil from the LP Separator to preheat the oil from the first stage separator. This would reduce the overall heat and electrical load.

Use the discharged hot water in a Rankine cycle to produce electricity. This would reduce the load on power generation reducing fuel gas usage.

Use some of the pressure drop across LCV 01 to drive a turbine to produce electricity. This would reduce the load on power generation reducing fuel gas usage.

[Question total: 20 marks]

**Q.59 Question 59
9: Energy efficiency**

a) An anti-surge control system is to be designed for a centrifugal compressor. Three options are under consideration:

- suction throttling at constant speed.
- discharge gas recycling at constant speed.
- speed control.

In all cases the compressor discharge pressure is fixed. With respect to energy requirements discuss the merits or otherwise of the three options. **[6 marks]**

Solution:

Surge occurs at low volume throughputs.

Suction throttling increases the suction gas volume and reduces suction pressure. For a fixed discharge pressure the compression ratio will increase thus the energy required by the compressor will also increase.[✓]

Gas recycling adds a portion of the discharge gas into the compressor suction. The mass of gas through the compressor will increase hence increasing compressor power.[✓]

At low gas volumes speed control allows the compressor to be operated at lower volumes without surging. Speed control would use less energy than the other two options as neither suction volume or mass is increased.[✓]

b) A refrigeration plant is to be designed using propane as the refrigerant. It is proposed to use a variable speed drive for the propane compressor to save energy during winter and summer operations. The mass of process gas is fixed and has to be cooled to -35°C at all times. The refrigeration system is a single stage design and refrigerant condensation is via an air cooler. Summer temperatures are 35°C and winter temperatures are 10°C. Use the attached propane pressure-enthalpy chart in Fig. 13 to determine the percentage energy saving from winter to summer operations that speed control provides.

Assume the refrigeration cycle is isentropic and the temperature approach in the air cooler and evaporator is 5°C.

Include your marked up pressure-enthalpy chart with your exam paper. Put your student number on the chart. **[10 marks]**

Solution:

Assuming a 5°C temp approach the propane will require to be compressed to a temperature of:

$$\text{Summer } 35+5 = 40^\circ\text{C}$$

$$\text{Winter } 10+5 = 15^\circ\text{C}$$

With the evaporator in both cases needing to reach -40°C.[✓]

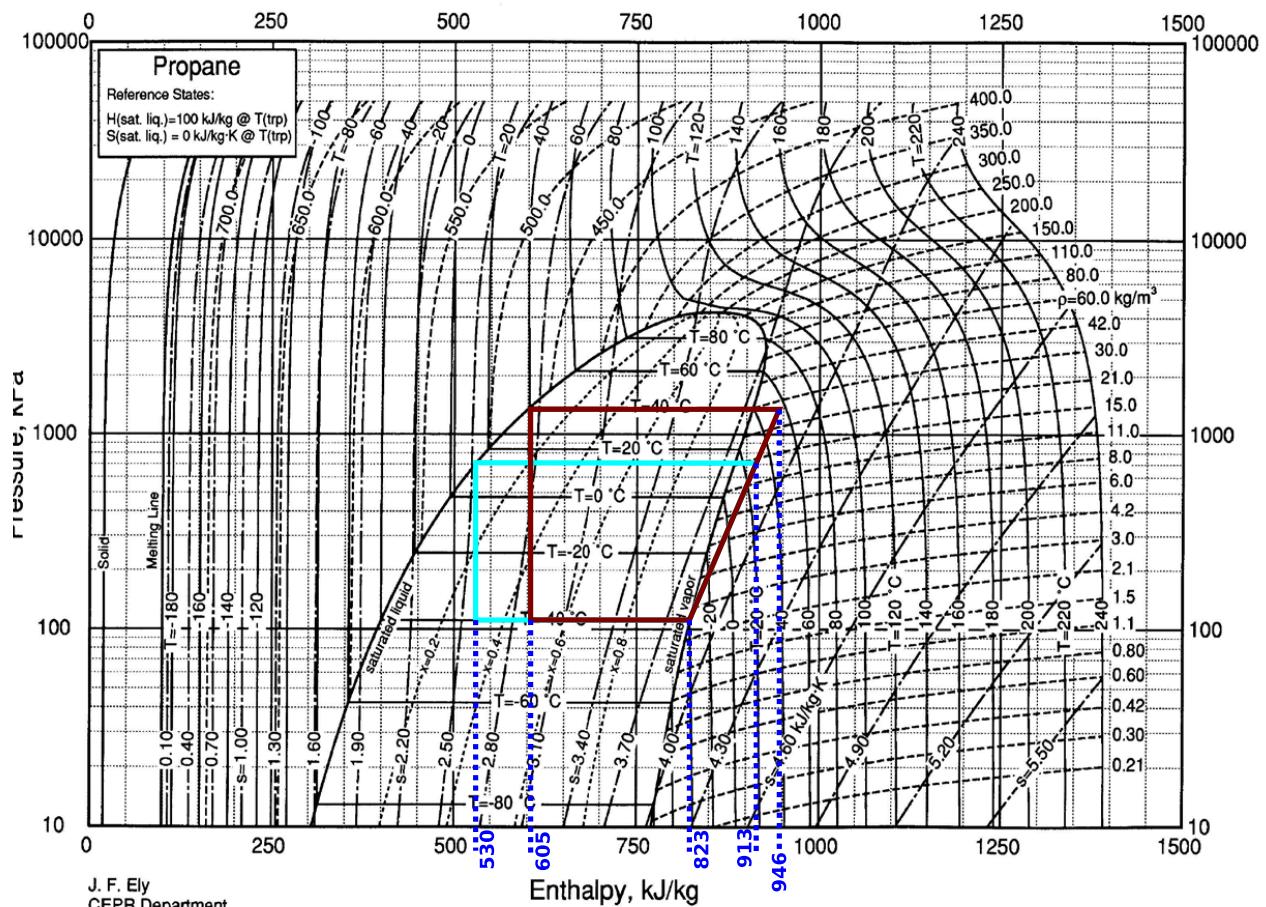
The student should sketch both winter and summer cycles onto P-H chart[✓]

| | Winter | Summer | Units |
|-------------|-------------------|-------------------|-------|
| Compression | $913 - 823 = 90$ | $946 - 823 = 123$ | kJ/kg |
| Evaporation | $823 - 530 = 293$ | $823 - 605 = 218$ | kJ/kg |

Thus in summer the unit has a refrigeration efficiency of $218/123 = 1.77$ (it removes 1.77 Joules per Joule of compression energy) and in winter its $293/90 = 3.26$.[✓]

Without speed control, the summer cycle must be used all year round, but with speed control only $1.77/3.26 = 54\%$ of the energy is required in winter, so a saving of 46%.[✓]

No marks to be deducted for slightly different numbers extracted from PH chart.



- c) Estimate the percentage speed reduction of the refrigeration compressor from summer to winter. [4 marks]

Solution:

From p -H chart summer compression pressure = 1400 kPa

From p -H chart winter compression pressure = 700 kPa

Precision is challenging with the log scale, so wide allowance for pressure estimates is given.

[2/4] Pressure is proportional to impeller tip speed V^2 ; hence, summer to winter speed change = $(700/1400)^{1/2} = 0.707$

[1/4] So % change in speed summer to winter, $\approx 30\%$ reduction

[Question total: 20 marks]

If you use this graph, you must attach it to your exam booklet using the provided tag.
 Student ID: _____

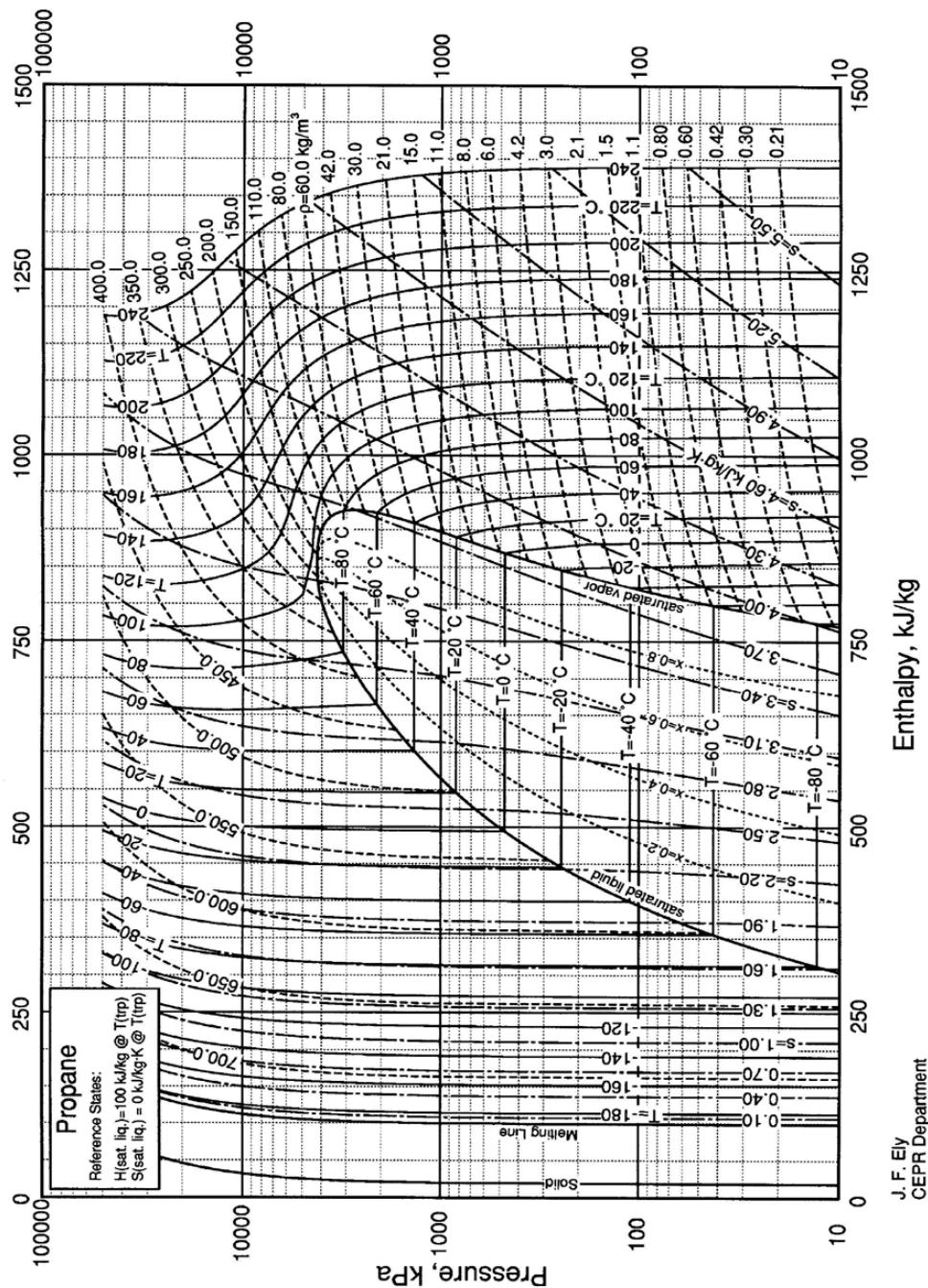


Figure 13: A Mollier chart for propane.

10 10: Water injection

Q.60

Question 60

Calculate the dissolved oxygen concentration of seawater at 5 °C and 20 °C.

[Question end]

Q.61**Question 61**

A seawater deaeration column utilising gas stripping in a single packed bed is operating with the following conditions.

Stripping gas is nitrogen. The oxygen partial pressure contained within the nitrogen at inlet and outlet conditions is 0 and 0.02 atm respectively. Seawater counter flows to the nitrogen. The seawater enters the column with an oxygen concentration of $8 \times 10^{-6} \text{ mol fraction}$, the seawater leaves the bottom of the column with a concentration of $0.05 \times 10^{-6} \text{ mol fraction}$. The column operates at 10° C . The height of the packing is 3.5 m.

The information is to be used to design a new column to achieve an oxygen outlet concentration of $0.04 \times 10^{-6} \text{ mol fraction}$. The other conditions for the new column are as above. Calculate the packing height which would be required in the new column.

Henry's constant for seawater can be taken as

| Temperature $^\circ \text{C}$ | $H_{O_2} \text{ atm / mol fraction}$ |
|-------------------------------|--------------------------------------|
| 0 | 22,500 |
| 5 | 29,100 |
| 10 | 32,700 |
| 15 | 36,400 |
| 20 | 40,100 |
| 25 | 43,800 |
| 30 | 47,500 |
| 35 | 50,70 |
| 40 | 53,500 |
| 45 | 56,300 |
| 50 | 58,800 |
| 60 | 62,900 |
| 70 | 66,300 |
| 80 | 68,700 |
| 90 | 69,900 |
| 100 | 70,100 |

[Question end]

Datasheet

Unit conversions

1 US barrel = 42 US gallons = 0.159 m³

1 ft = 12 in = 0.3048 m

1 lb = 0.453592 kg

°F = (°C × 9/5) + 32

°C = (°F – 32) 5/9

Physical constants/properties

Air molecular weight = 29 g mol⁻¹

Darcy's equation

$$\Delta p = \frac{\rho f L v^2}{2 D}$$

Blasius correlation (turbulent flow)

$$f = 0.316 \text{ Re}^{-0.25}$$

| Pipe fitting | Equiv. length (L/D) |
|------------------------------------|---------------------|
| 90° standard elbow | 30 |
| 45° standard elbow | 16 |
| 90° long elbow | 20 |
| 90° street elbow | 50 |
| 45° street elbow | 26 |
| Square corner elbow | 57 |
| Standard tee (flow through run) | 20 |
| Standard tee (flow through branch) | 60 |
| Close pattern return bend | 50 |

Table 1: A list of equivalent lengths for single phase fluid flow calculations.