

This document contains the tutorial questions for the 15-credit EG5099 Upstream Oil & Gas Processing 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. 8 to Q. 18.
- Tutorial 3 Separation: Q. 20
- Tutorial 4 Produced Water: Q. 21 to Q. 23.
- Tutorial 5 Compression: Q. 24 to Q. 26.
- Tutorial 6 Gas Treatment: Q. 27 to Q. 30.
- Tutorial 7 Multi-phase: Q. 33 to Q. 37.
- Tutorial 8 Hydrates: Q. 38 to Q. 40.
- Tutorial 9 Energy Efficiency: Q. 41 to Q. 50.
- Tutorial 10 Water Injection: Q. 51 to Q. 52.

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 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  $\alpha$ :

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

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  $\alpha$

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

Calculate  $A$  and  $B$ :

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

$$B = 0.08664 \left( \frac{P}{0.95877 P_c} \right) = 0.001871P \quad (25)$$

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.

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

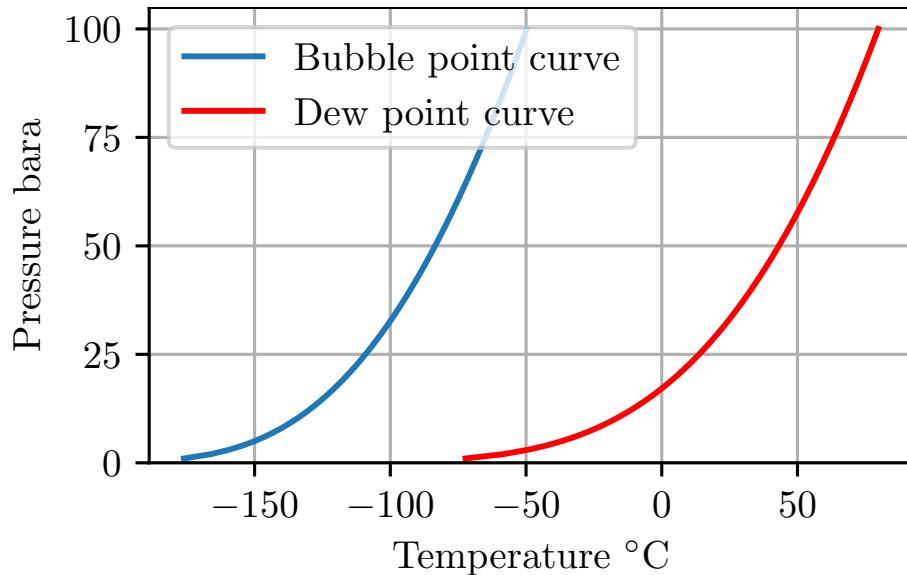
where  $P$  is the absolute pressure,  $P_{ci}$  is the critical pressure of component  $i$  in the same units as  $P$ ,  $T$  is the absolute temperature,  $T_{ci}$  is the critical temperature of component  $i$  in the same units as  $T$ , and  $\omega_i$  is the acentric factor of component  $i$ .

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:

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

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

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

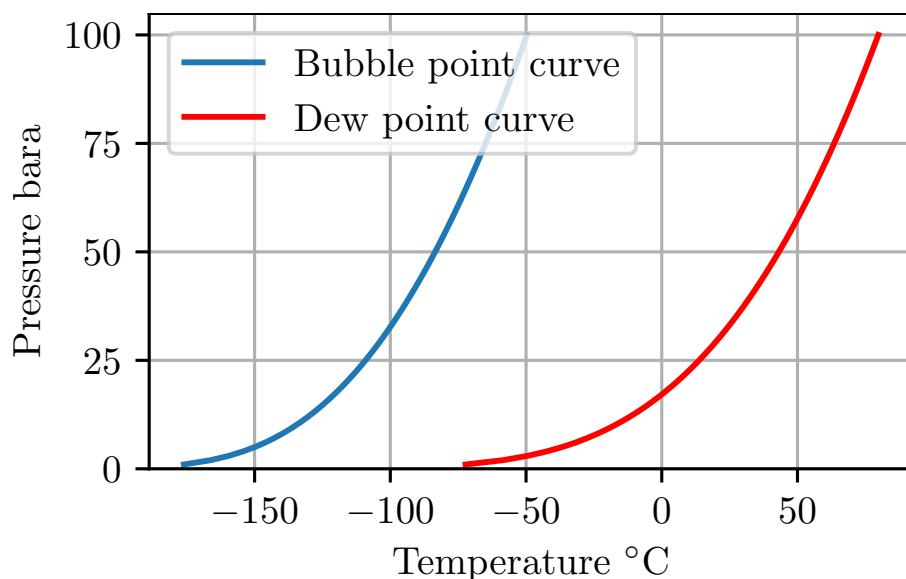
$$K_{i-butane} = 9.744 \times 10^{-8} \quad (39)$$

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

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.

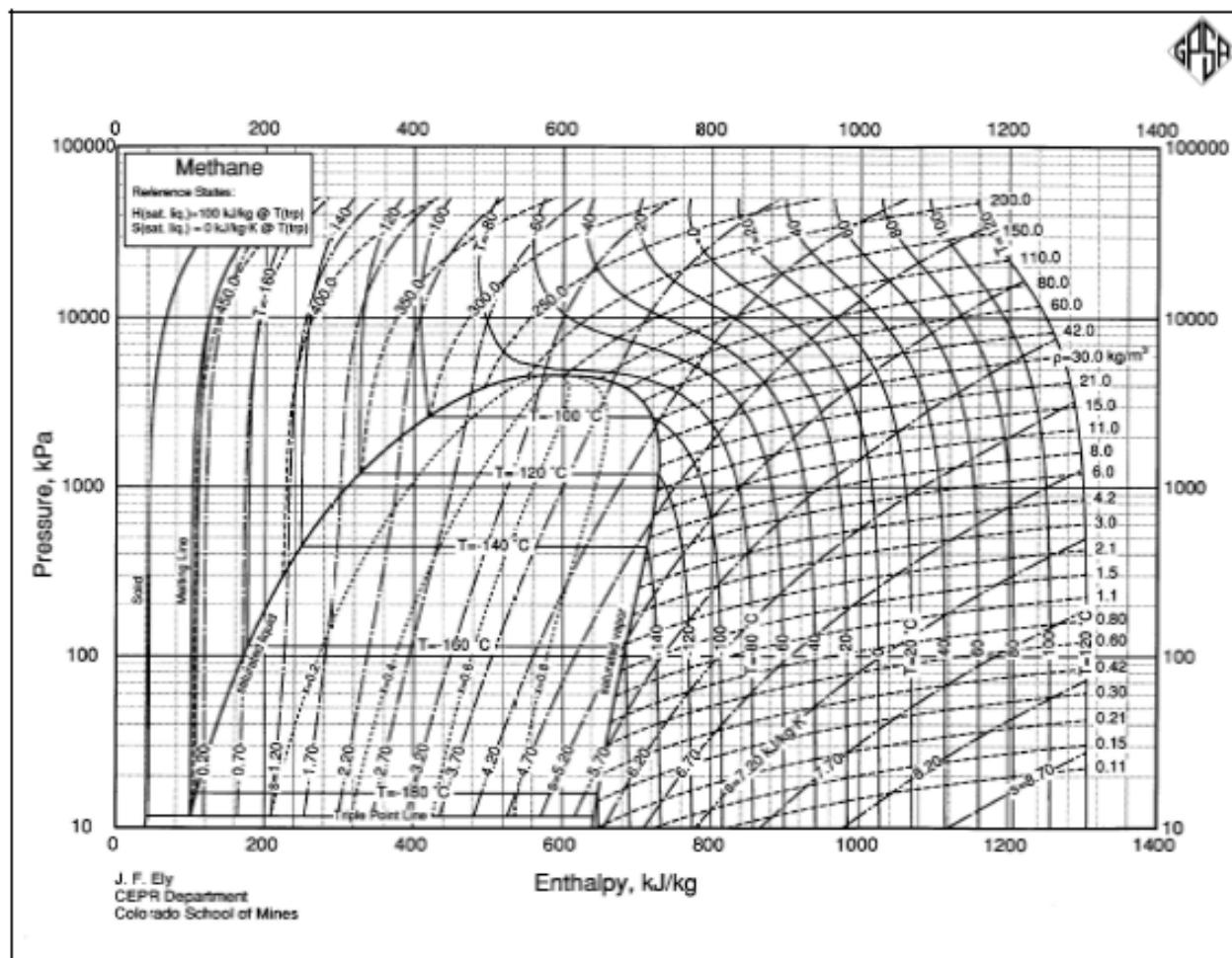
**[Question end]**

#### 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]

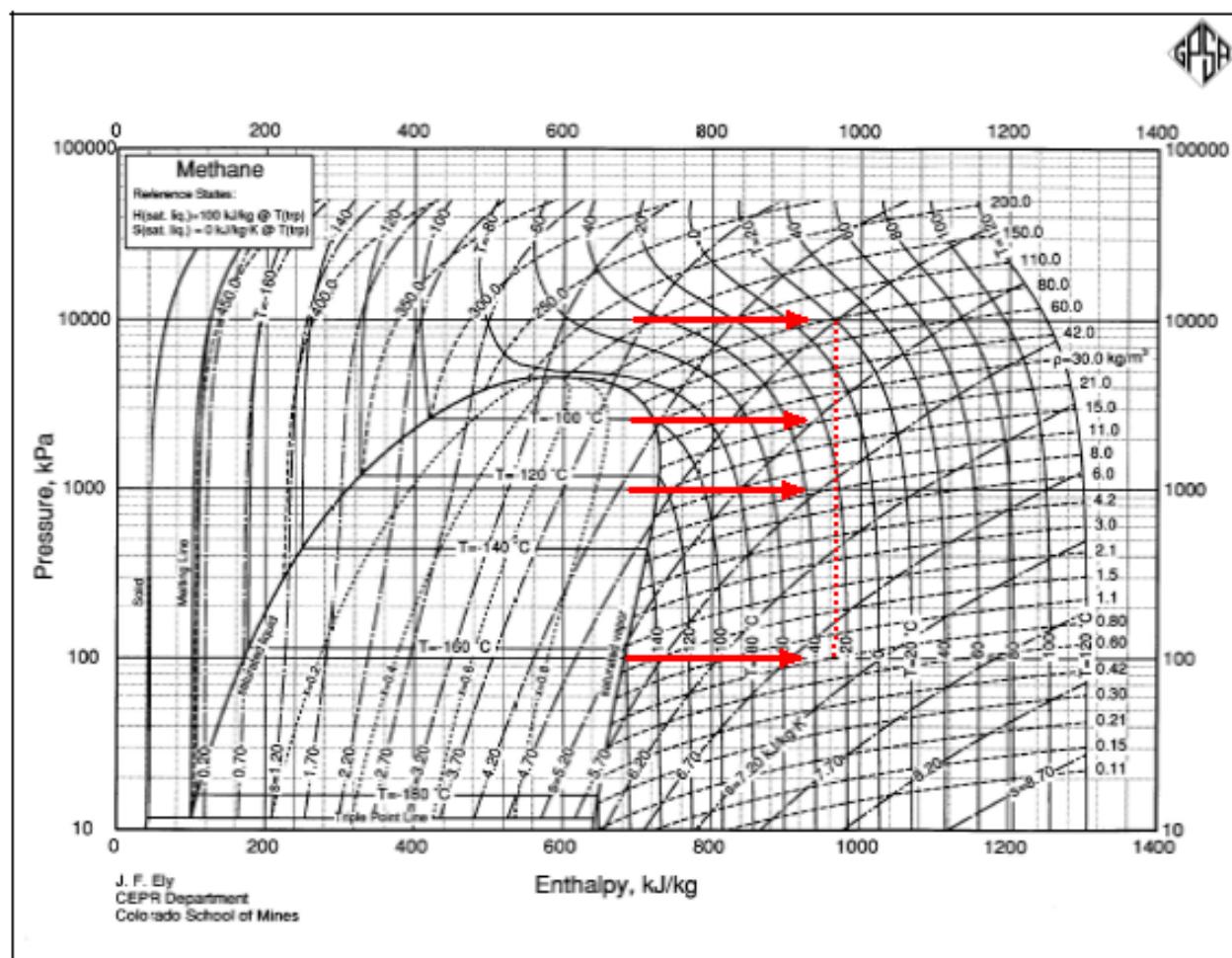
As the pipeline fills with methane the back pressure increases. Calculate the choke downstream temperature at back pressures of 10 and 25 bara. [-20 °C, -18 °C]



## Solution:

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

Find the enthalpy of the methane using the initial pressure  $P_0 = 100$  bara and temperature  $T_0 = 20^\circ\text{C}$  and find the new temperature when expanded to  $P = 1$  bara / 10 bara / 25 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 <sub>2</sub>	0.0046
CO <sub>2</sub>	0.0030
H <sub>2</sub> 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<sup>-3</sup>]

**Solution:**

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

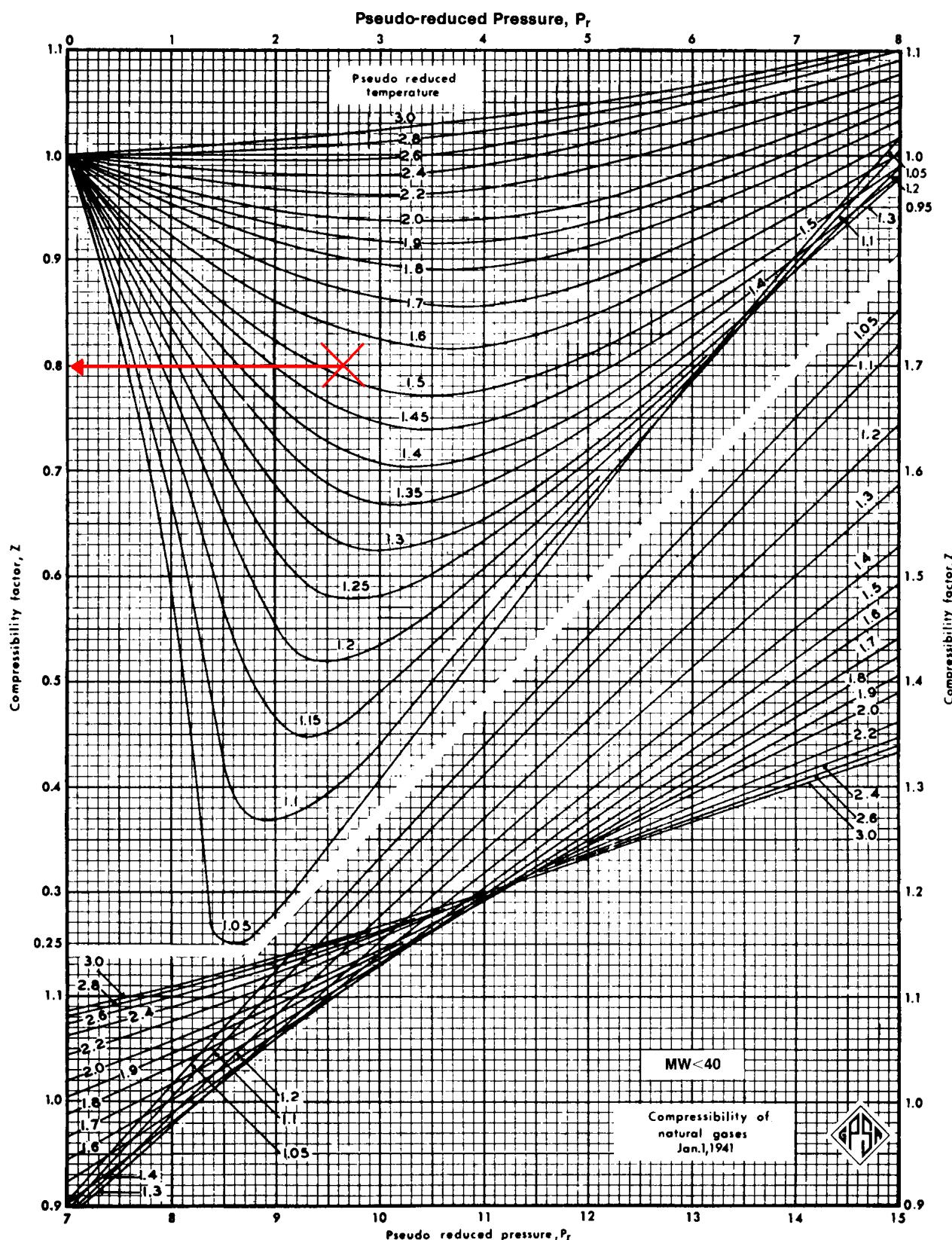
Component	$y_i$	$P_C$	$P_C \cdot y_i$	$T_C$	$T_C \cdot y_i$	$M$	$M \cdot y_i$
N <sub>2</sub>	0.0046	3.40	0.01564	126	0.6	28.02	0.13
CO <sub>2</sub>	0.003	7.38	0.02214	304	0.9	44.01	0.13
H <sub>2</sub> S	0.1438	8.96	1.28845	373	53.6	34.08	4.90
Methane	0.8414	4.60	3.87044	191	160.7	16.04	13.50
Ethane	0.0059	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

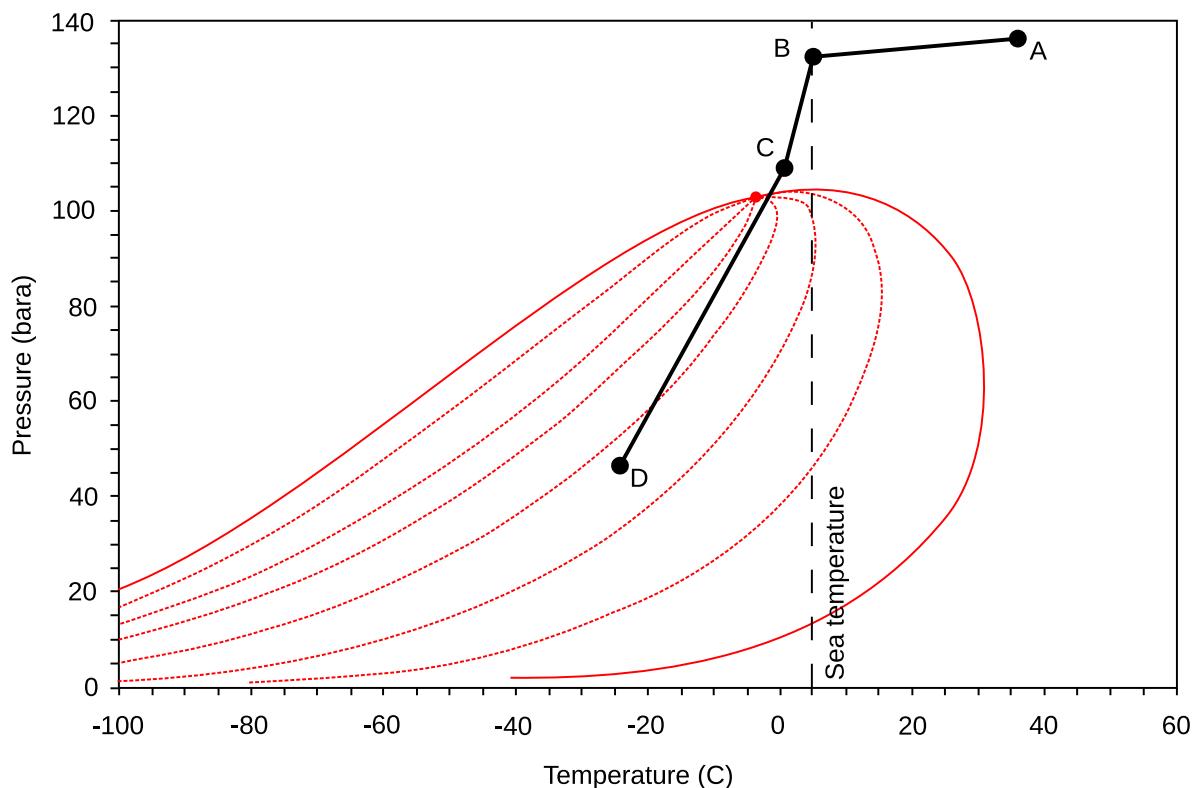
**Question 6**  
**Past exam question**


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

- 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.  $\checkmark$

[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.  $\checkmark$

[5/5]

**[Question total: 20 marks]**

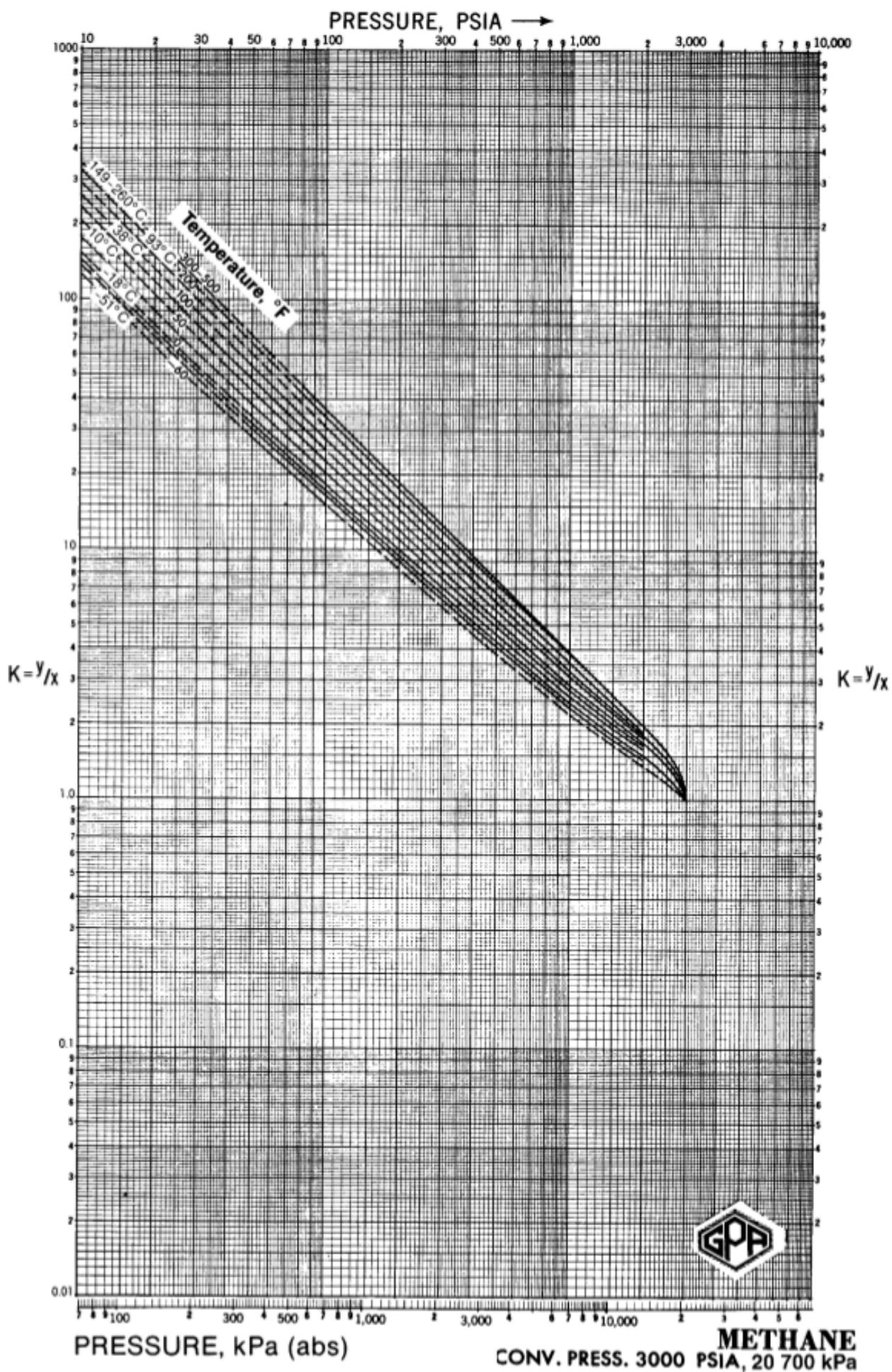
#### Q.7

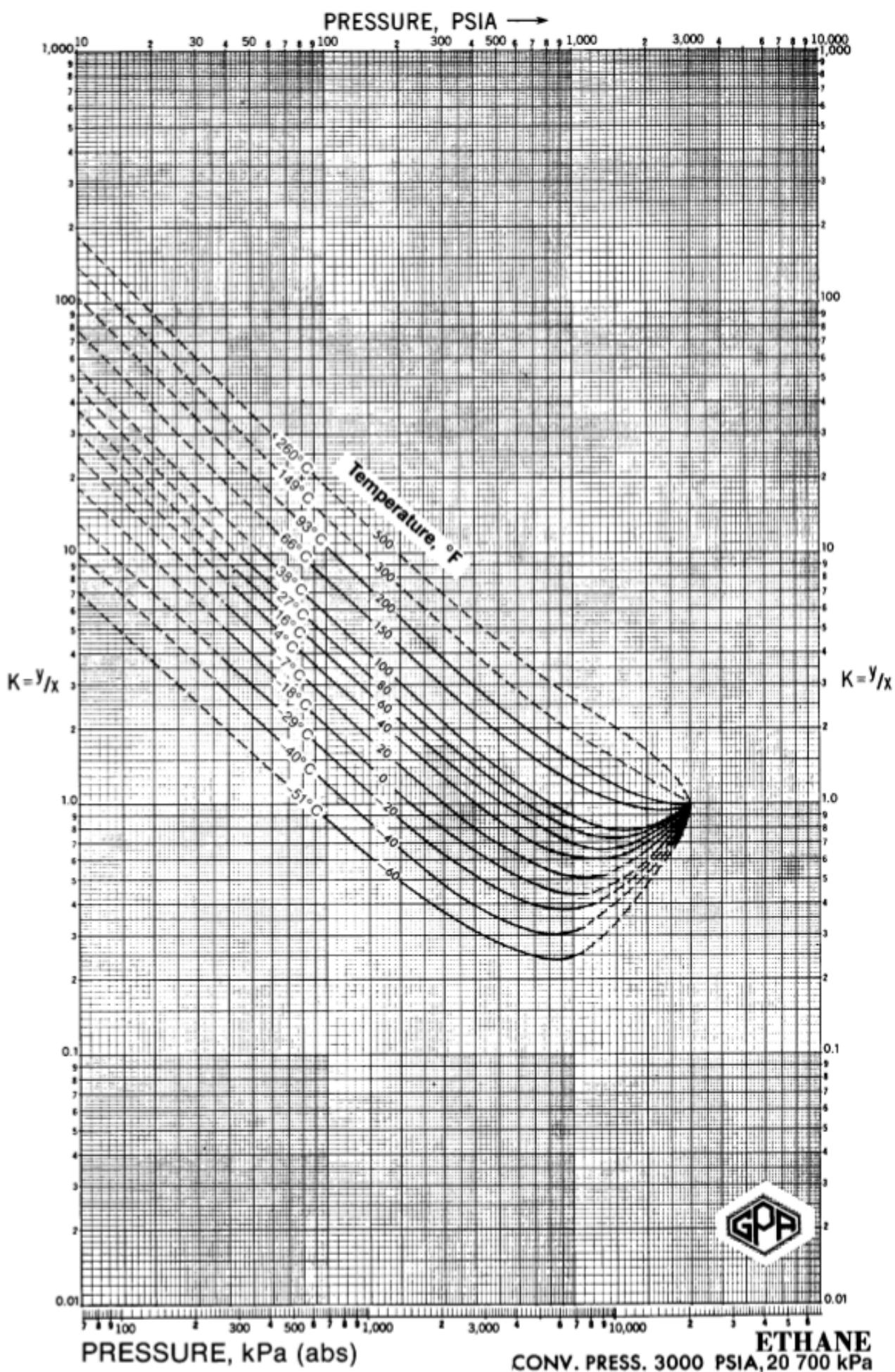
#### Question 7

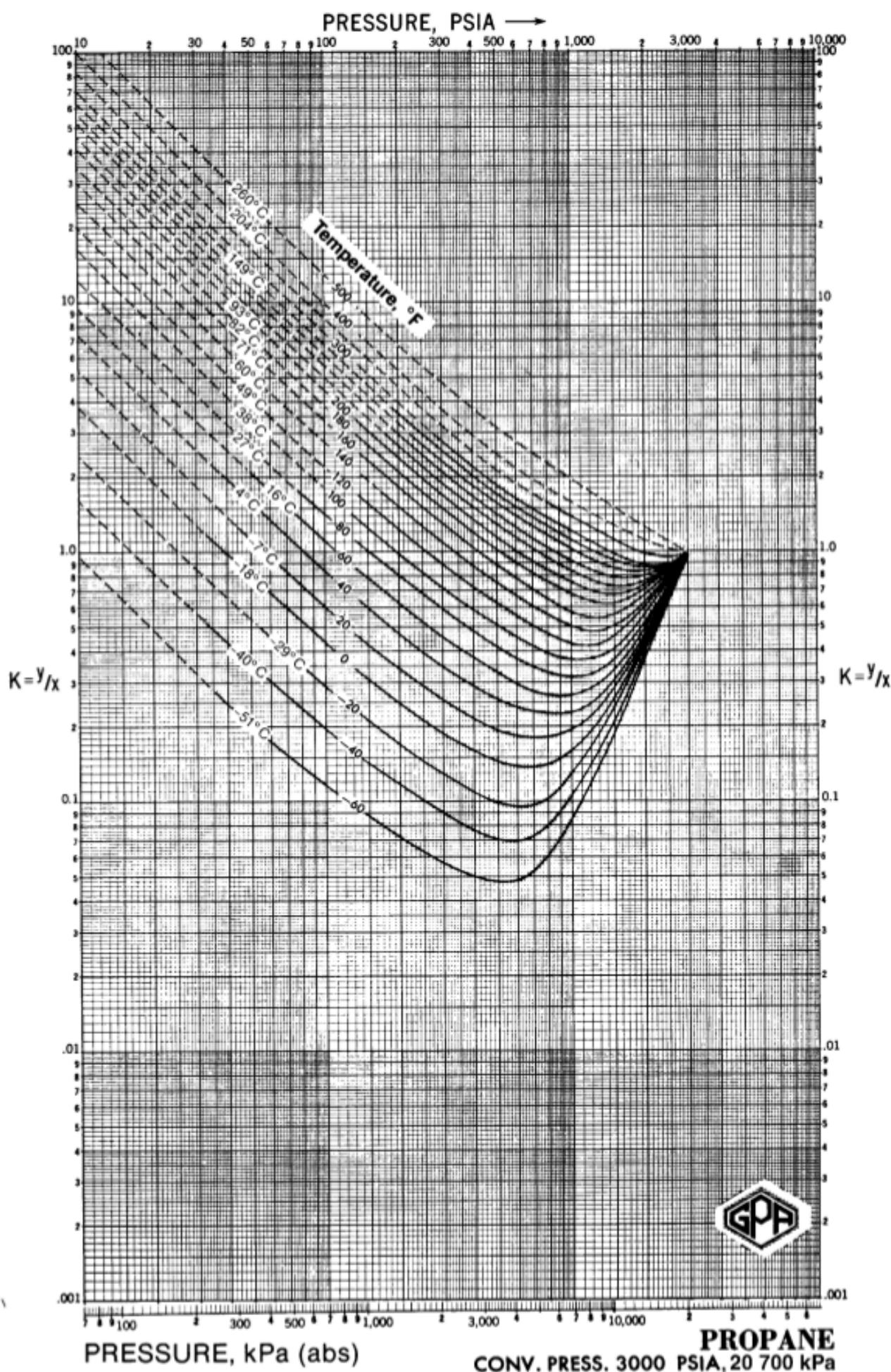
The following mixture flows in a subsea pipeline. The inlet temperature and pressure of the pipeline are 150 °F and 1000 psia the outlet conditions are 100 °F and 200 psia. Calculate the vapour liquid ratio at the pipeline inlet and outlet using the attached K value charts.

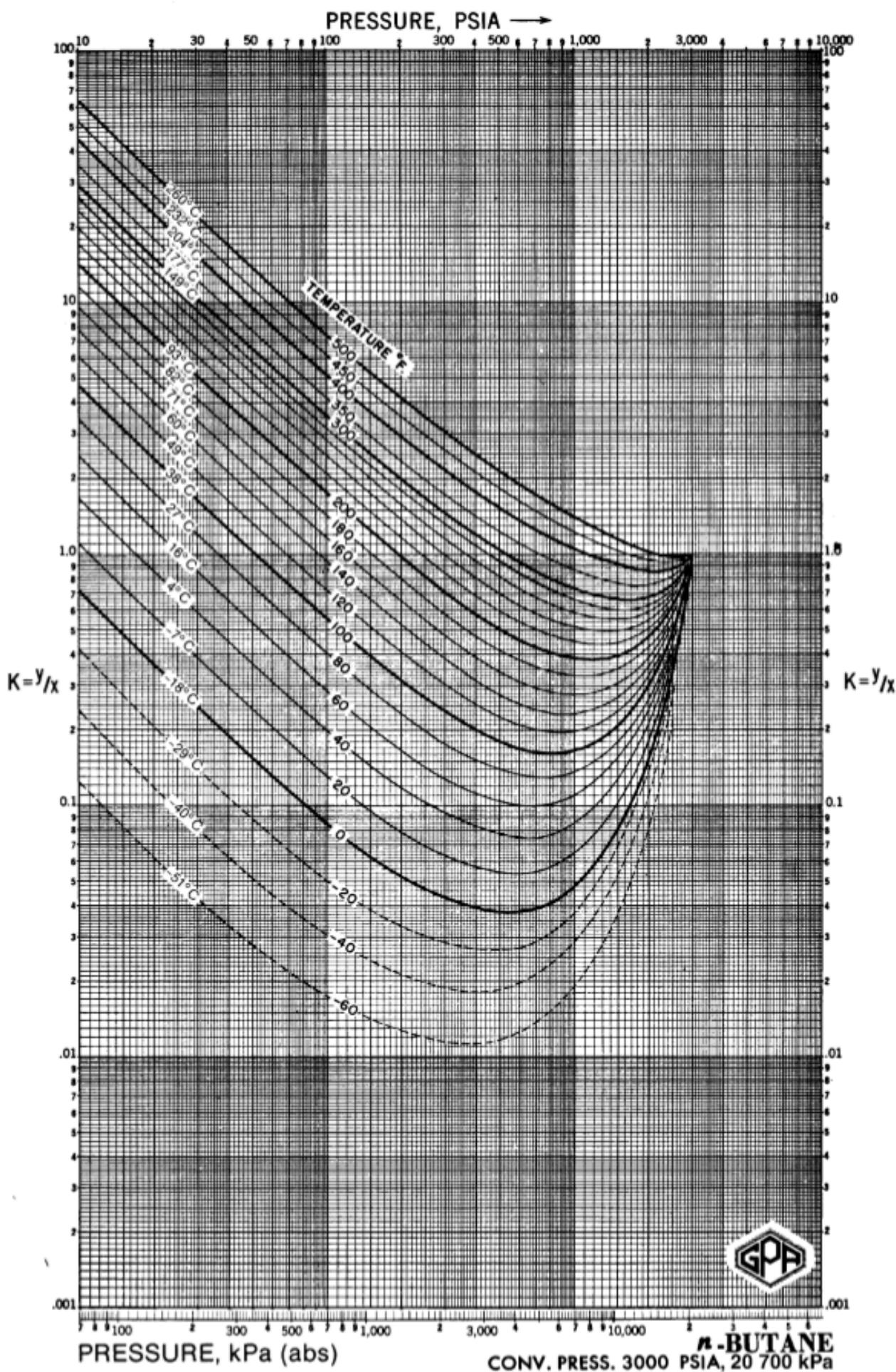
[Inlet V=0.14, L=0.86, Outlet V=0.67, L=0.33]

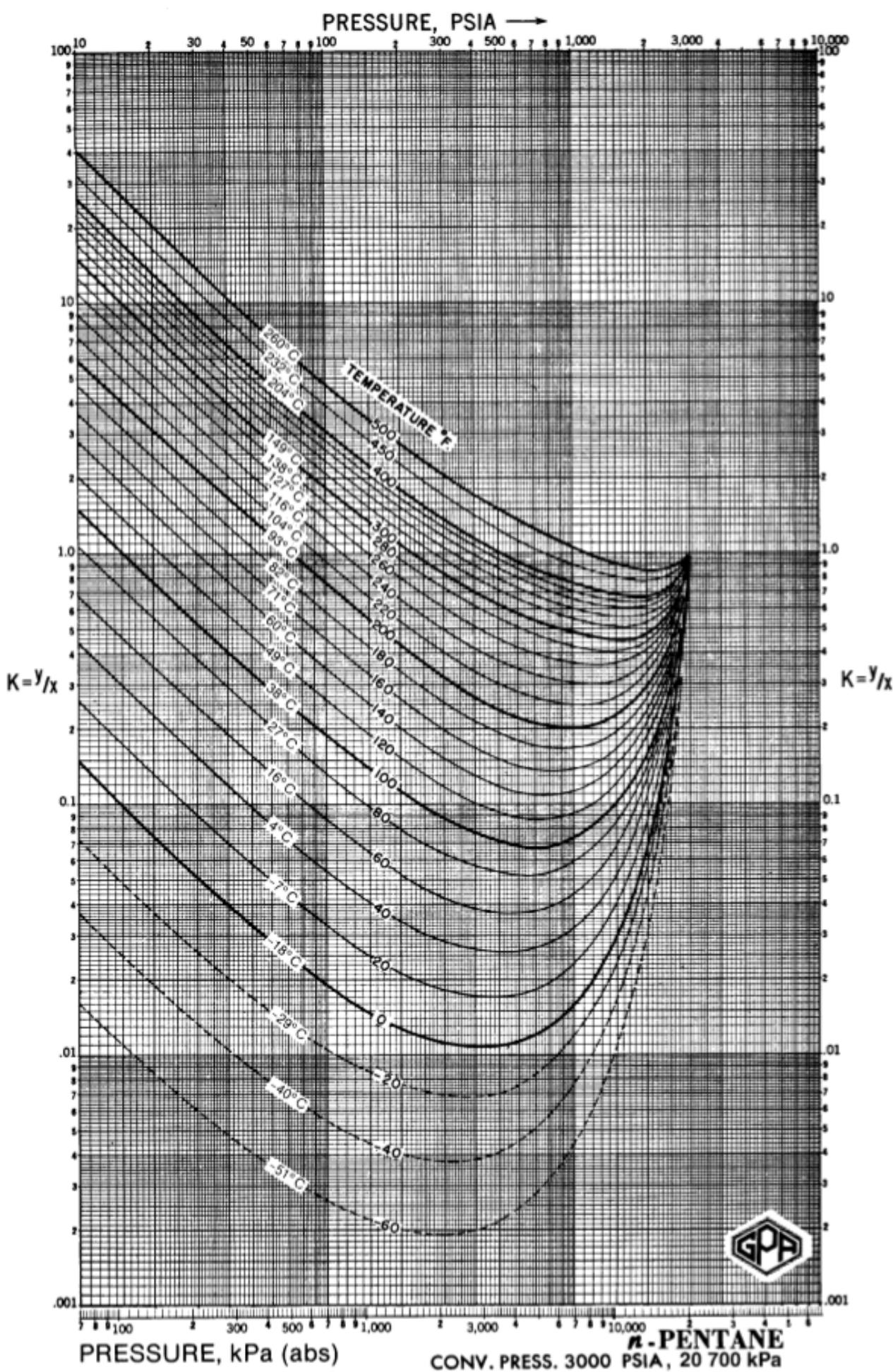
Component	Mol fraction
Methane	0.2
Ethane	0.3
Propane	0.15
n-Butane	0.1
n-Pentane	0.1
n-Octane	0.15

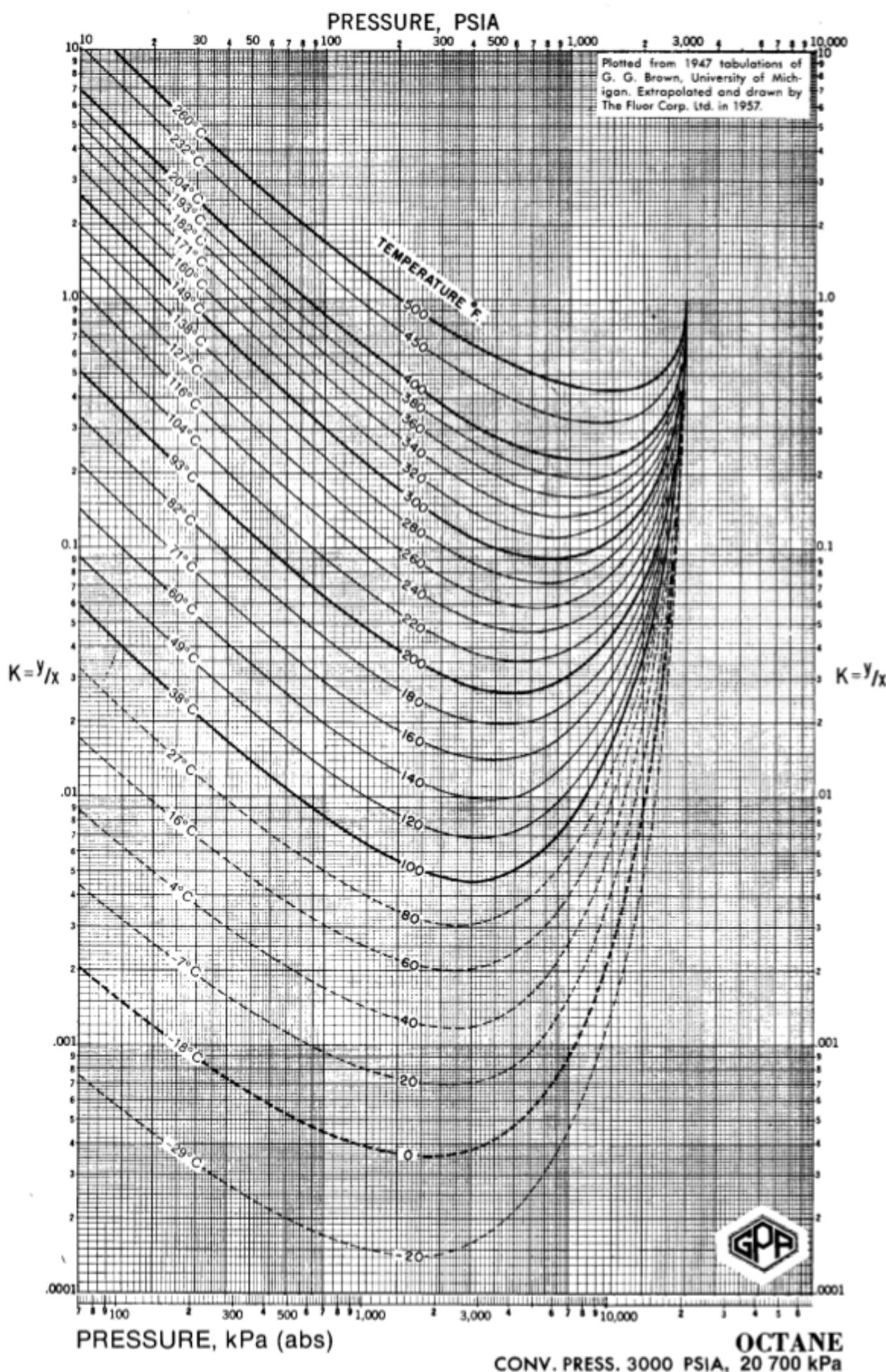












**Solution:**

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]

## 2 Fluid flow

### Q.8

#### Question 8

An oil which has a viscosity of 50 cP ( $0.05 \text{ kg m}^{-1} \text{ s}^{-1}$ ) and a density of  $50 \text{ lb ft}^{-3}$  ( $801.3 \text{ kg m}^{-3}$ ) 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.9

#### Question 9

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.10****Question 10**

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.11****Question 11**

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<sup>-1</sup>]

**Solution:**

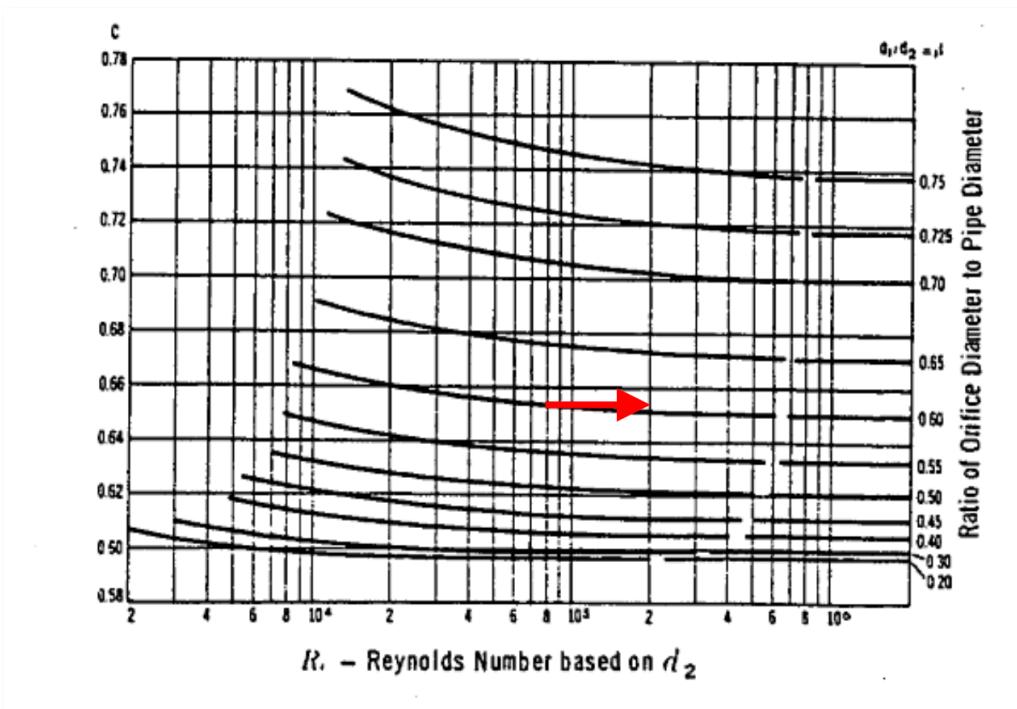
We will calculate the flowrate through the orifice using the following equation (from lecture notes)

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

Firstly the pressure gradient from the mercury reading is calculated (converting mm Hg to Pa: 0.75 mm Hg = 10.2 mm H<sub>2</sub>O = 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.  $\text{Re} \gg 2000$ )



$C = 0.62$ , 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

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

The assumption of turbulent flow was valid

**[Question end]**

### Q.12 Question 12

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:**

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

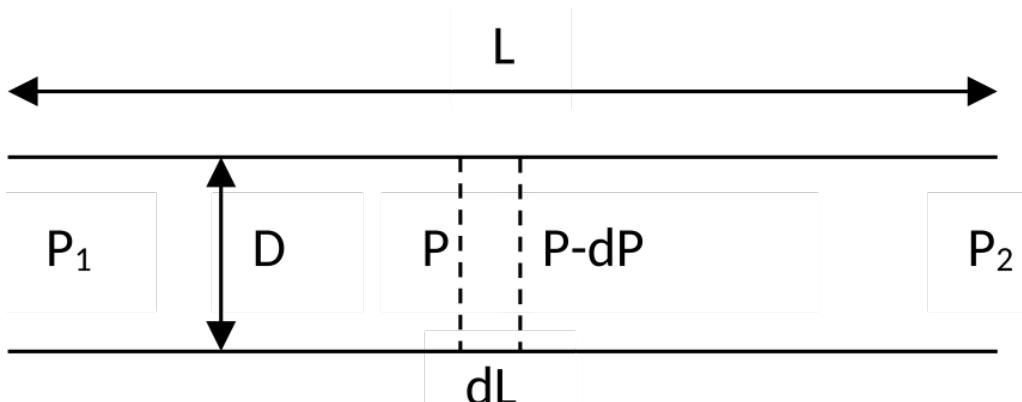
The pressure drop is given as

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

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  $\rho$  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  $p = p_2$

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

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

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

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

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

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

### [Question end]

#### Q.13 Question 13

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<sup>-1</sup>, 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  $\rho_1$ .

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

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  $V$  (m<sup>3</sup>)

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 <sup>3</sup>	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 <sup>2</sup>	liter	0.084 784	lb mol	°R	hp-hr	0.000 780 48	
gm mol	K	kPa	m <sup>3</sup>	0.0083145	lb mol	°R	Kw-hr	0.000 582 00	
lb mol	°R	atm	ft <sup>3</sup>	0.730 24	lb mol	°R	ft-lb	1 545.3	
lb mol	°R	in.Hg	ft <sup>3</sup>	21.850					
lb mol	°R	mm Hg	ft <sup>3</sup>	554.98	k mol	K	joule	8 314.5	
lb mol	°R	lb/in <sup>2</sup>	ft <sup>3</sup>	10.732					
lb mol	°R	lb/ft <sup>2</sup>	ft <sup>3</sup>	11 545.3					
lb mol	K	atm	ft <sup>3</sup>	1.3144					
lb mol	K	mm Hg	ft <sup>3</sup>	998.97					
k mol	K	kPa	m <sup>3</sup>	8.3145					
k mol	K	bar	m <sup>3</sup>	0.083 145					

$$\frac{n}{V} = \frac{p_1}{RT_1Z} = \frac{2000}{0.0083415 \cdot 283 \cdot 0.88} \quad (84)$$

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

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.14 Question 14

What is the riser base pressure if the riser from Q. 13 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. 13 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.

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

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

**[Question end]**

**Q.15 Question 15**

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 m<sup>3</sup> min<sup>-1</sup>. Neglect frictional pressure drop. [1.575 bara]

**Solution:**

Use Bernoulli's equation

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

(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.16 Question 16**

Calculate the speed of sound in air at 15 °C. [340 m s<sup>-1</sup>]

Calculate the speed of sound in water. [1466 m s<sup>-1</sup>]

**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<sup>-1</sup>, 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

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

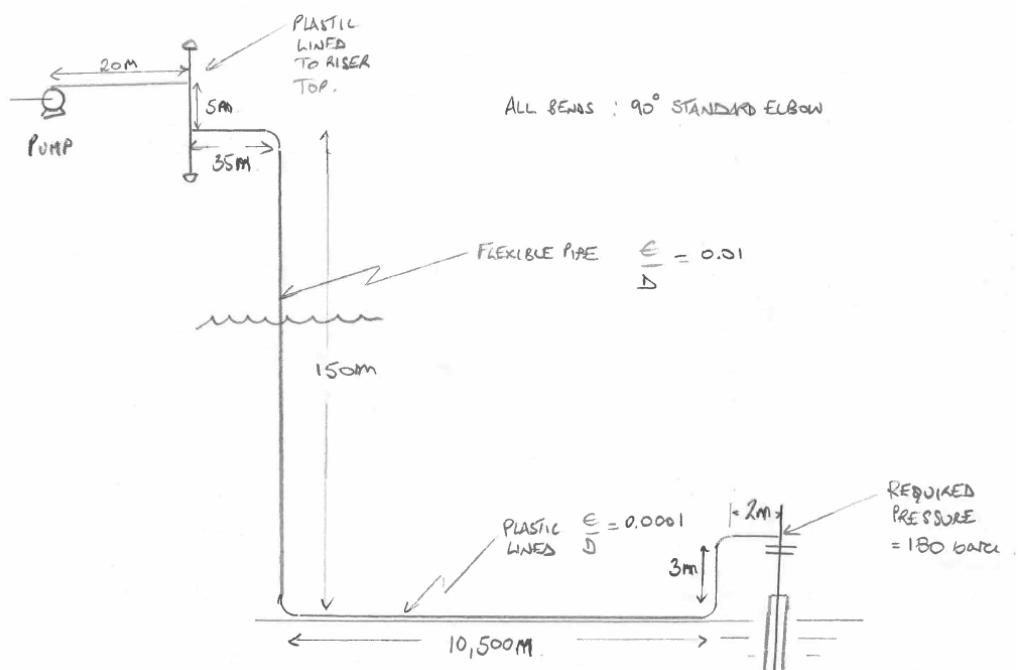
where the value for  $k_{\text{water}}$  is taken from the lecture notes.

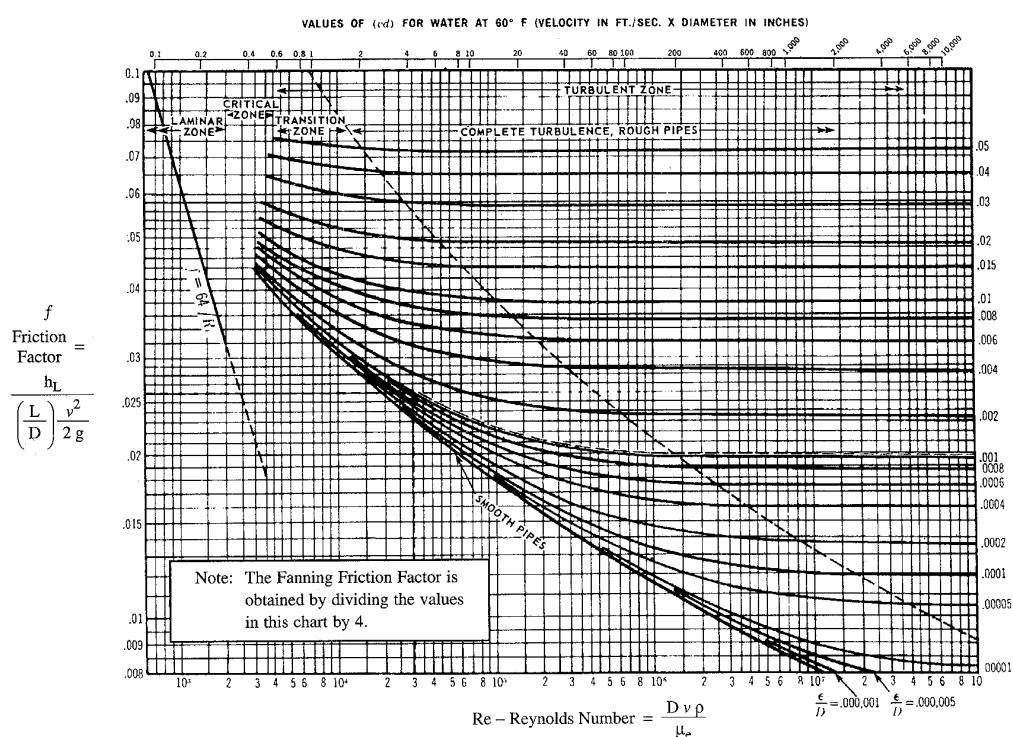
$$c = \sqrt{\frac{k}{\rho}} = \sqrt{\frac{2.15 \times 10^9}{1000}} = 1466.25 \text{ m s}^{-1} \quad (106)$$

**[Question end]**

**Q.17****Question 17**

A plastic lined, subsea water injection flowline is required to deliver 80,000 bbl day<sup>-1</sup> 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 bbl = 1 m<sup>3</sup>. Viscosity and density of sea water at 10 °C is 1.31 cP and 1020 kg m<sup>-3</sup> 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 $60^\circ$ from run of pipe Stem $45^\circ$ 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

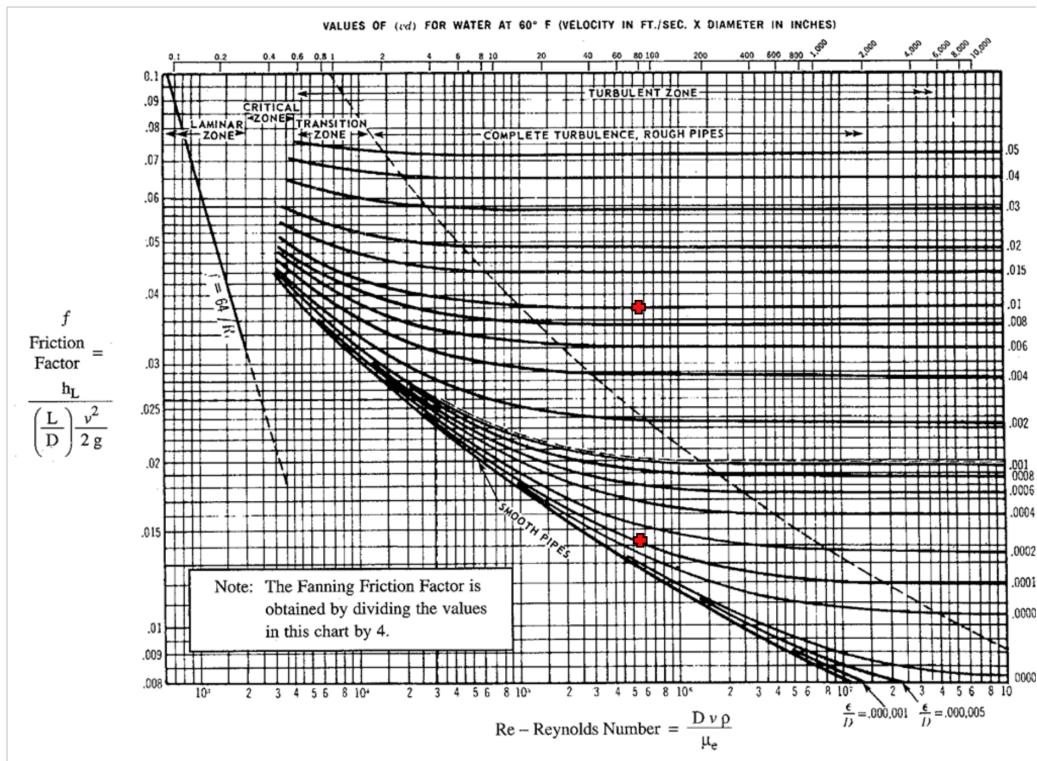
$$L = \underbrace{20 + 5 + 35 + 10500 + 3 + 2}_{\text{pipe length excluding riser}} + \underbrace{32.8 + 32.8}_{\text{equivalent length of fittings}} \quad (110)$$

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

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  $\mu u = 1.31 \text{ cP} = 0.00131 \text{ Pa s}$  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

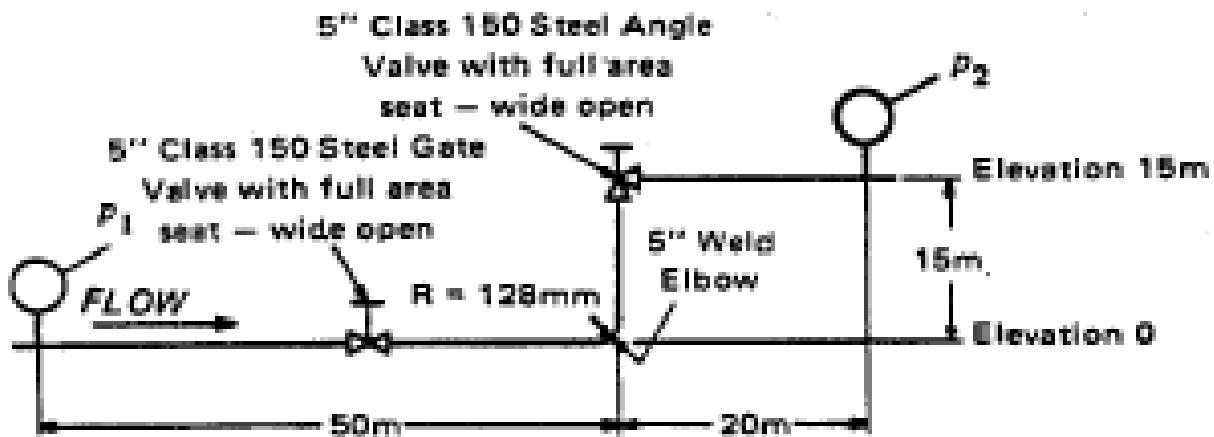
$$p_{pump} = 180 + 18.6 + 0.68 - 15.2 = 184.1 \text{ bar} \quad (116)$$

**[Question end]**

## Q.18

**Question 18**

For the following piping system has an oil flow rate of  $2300 \text{ l min}^{-1}$ , pipe diameter of 128.2 mm, liquid density of  $899 \text{ kg m}^{-3}$  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

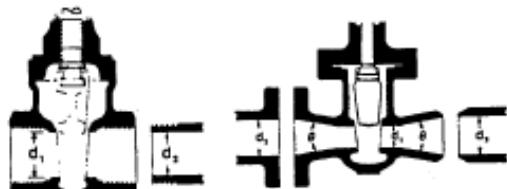
## Representative Resistance Coefficients (K) for Valves and Fittings

(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



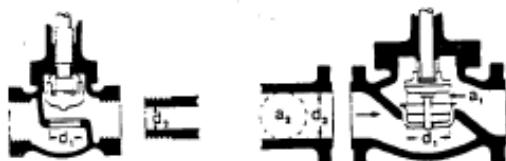
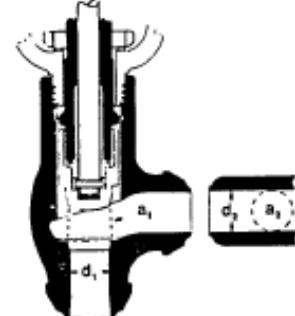
If:  $\beta = 1, \theta = 0 \dots K_1 = 8 f_T$   
 $\beta < 1 \text{ and } \theta < 45^\circ \dots K_1 = \text{Formula 5}$   
 $\beta < 1 \text{ and } 45^\circ < \theta < 180^\circ \dots K_1 = \text{Formula 6}$

## 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

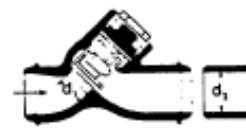
If:  $\beta = 1 \dots K_1 = 340 f_T$ If:  $\beta = 1 \dots K_1 = 55 f_T$ If:  $\beta = 1 \dots K_1 = 150 f_T$     If:  $\beta = 1 \dots K_1 = 55 f_T$ All globe and angle valves,  
whether reduced seat or throttled,If:  $\beta < 1 \dots K_1 = \text{Formula 7}$ 

## 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

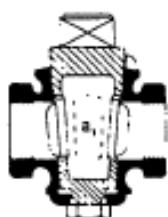
## "K" FACTOR TABLE - SHEET 4 of 4

## Representative Resistance Coefficients (K) for Valves and Fittings

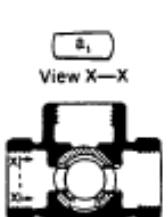
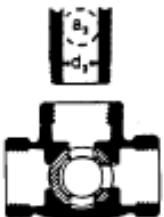
(for formulas and friction data, see page A-26)  
("K" is based on use of schedule pipe as listed on page 2-10)

## 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°

 $K = 30 f_T$ 

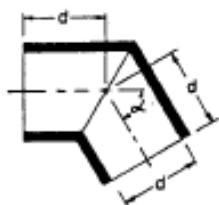
## 45°

 $K = 16 f_T$ 

## STANDARD TEES

Flow thru run . . . . .  $K = 20 f_T$ Flow thru branch . . . . .  $K = 60 f_T$ 

## MITRE BENDS



$\alpha$	$K$
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

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

$r/d$	$K$	$r/d$	$K$
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)

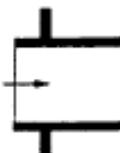
$$K_B = 31 (0.25 \pi \times 0.013 \times 3 + 0.5 \times 0.156) + 0.156$$

## CLOSE PATTERN RETURN BENDS

 $K = 50 f_T$ 

## PIPE ENTRANCE

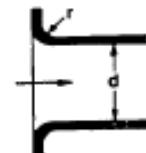
## Inward Projecting

 $K = 0.78$ 

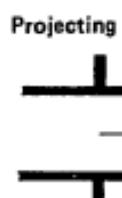
$r/d$	$K$
0.00*	0.5
0.02	0.28
0.04	0.24
0.06	0.15
0.10	0.09
0.15 & up	0.04

\*Sharp-edged

## Flush

For  $K$ ,  
see table

## PIPE EXIT

 $K = 1.0$  $K = 1.0$  $K = 1.0$

**"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  $\beta$  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

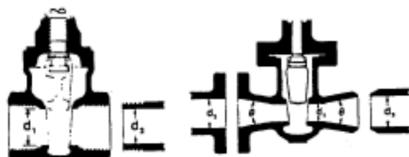
**Representative Resistance Coefficients (K) for Valves and Fittings**

(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



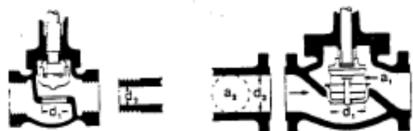
If:  $\beta = 1, \theta = 0 \dots K_1 = 8 f_T$   
 $\beta < 1 \text{ and } \theta < 45^\circ \dots K_1 = \text{Formula 5}$   
 $\beta < 1 \text{ and } 45^\circ < \theta < 180^\circ \dots K_1 = \text{Formula 6}$

**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  $= 120\sqrt{V}$

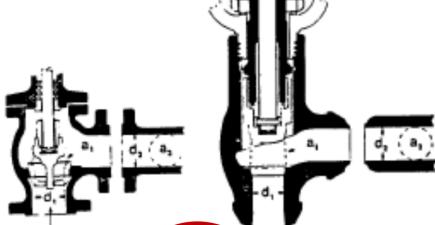
**GLOBE AND ANGLE VALVES**



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  
 $= 170\beta^2 \sqrt{V}$

**TILTING 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

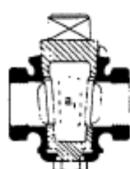
**Representative Resistance Coefficients (K) for Valves and Fittings**

(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**

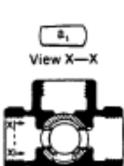
**Straight-Way**



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

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

**3-Way**



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

**STANDARD ELBOWS**

**90°**



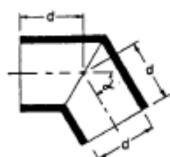
$K = 30 f_T$

**45°**



$K = 16 f_T$

**MITRE BENDS**



$\alpha$	$K$
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)

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

**CLOSE PATTERN RETURN BENDS**



$K = 50 f_T$

**PIPE ENTRANCE**

**Inward Projecting**

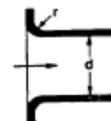


$K = 0.78$

$r/d$	$K$
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**



$K = 1.0$

**[Question end]**

**Q.19**

**Question 19**

**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 ?? of the datasheet. [10 marks]

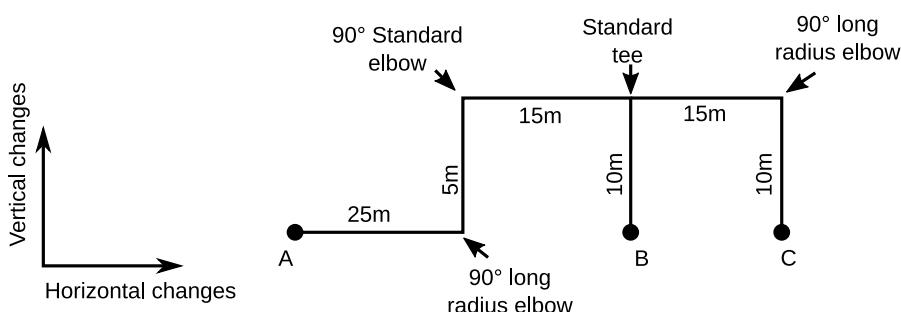


Figure 2: The flow diagram for the water injection system, and the pipeline. Please note the difference between vertical height changes and horizontal distance changes.

**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.

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]

✓

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

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

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:**

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.

[4/4]

[Question total: 20 marks]

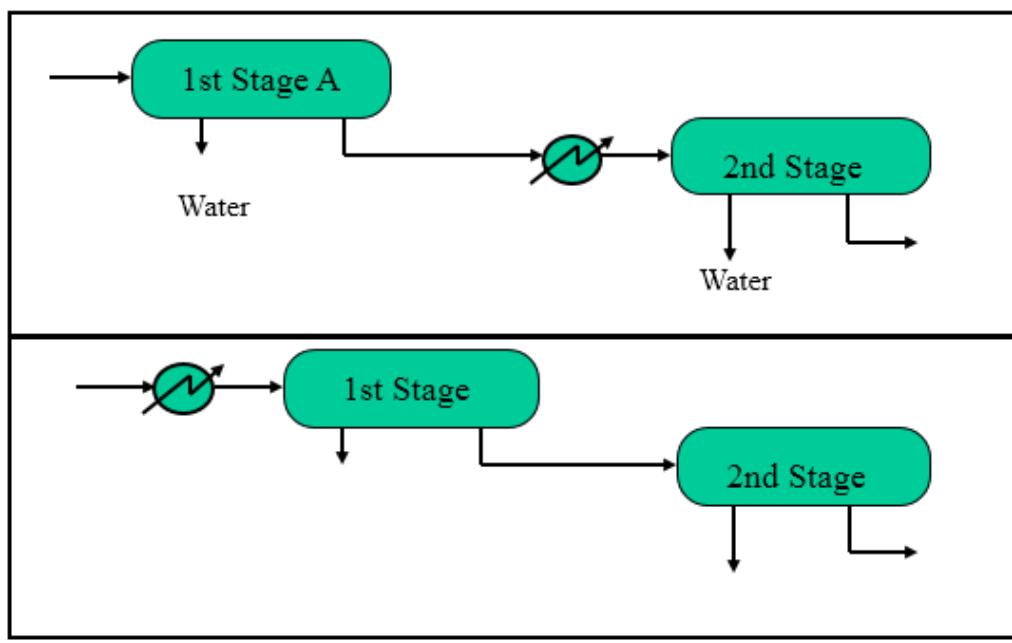
### 3 Separator

Q.20

#### Question 20

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  $\approx 3$ .
- 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

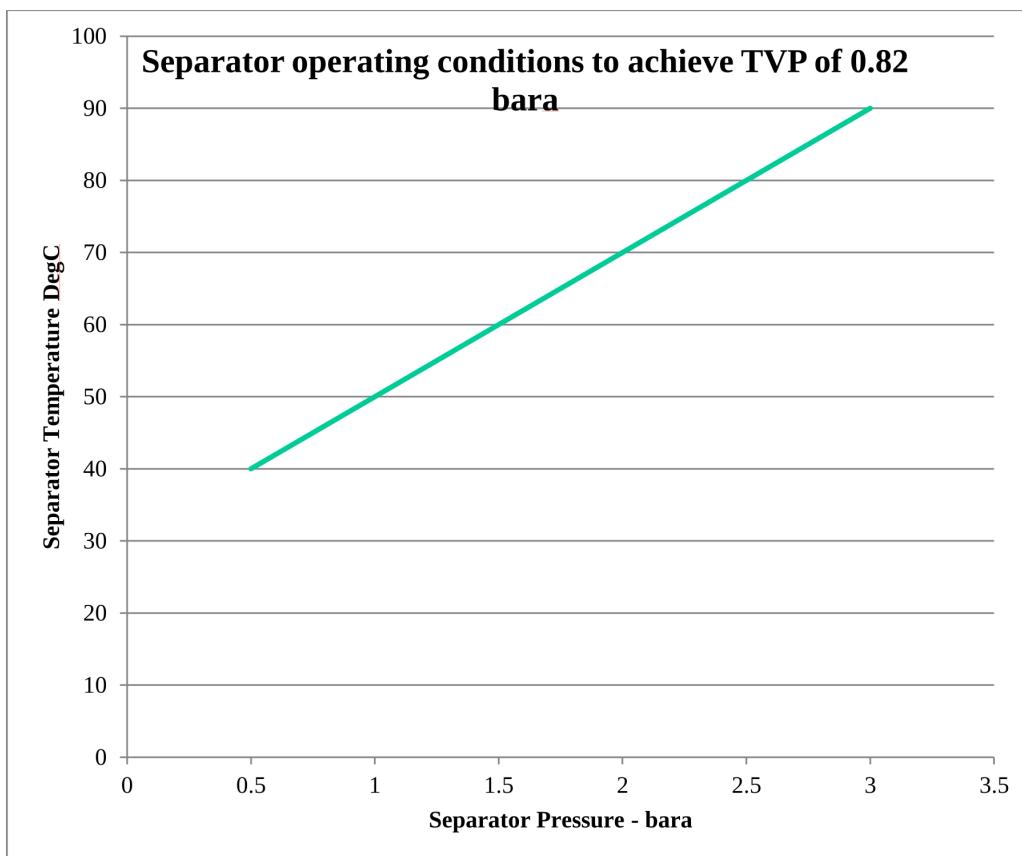
	Flowrate bpd	Specific gravity	Specific heat capacity kJ kg <sup>-1</sup> K <sup>-1</sup>
Oil	89,000	0.83	2.1
Water	60,000	1.025	4.2

## Oil specification

TVP = 0.82 bar at 37.8 °C

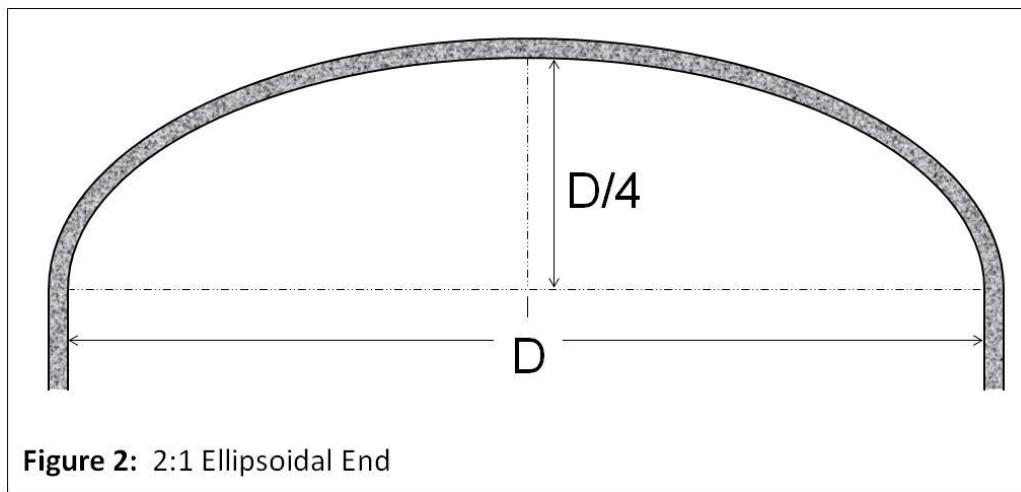
BS&W = 0.5 % v/v

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<sup>-2</sup> (normally 155 N mm<sup>-2</sup> for carbon steel),  $E$  is the joint efficiency (assume  $E = 1$ ), and  $D$  is the diameter in m.

The surface area is given by

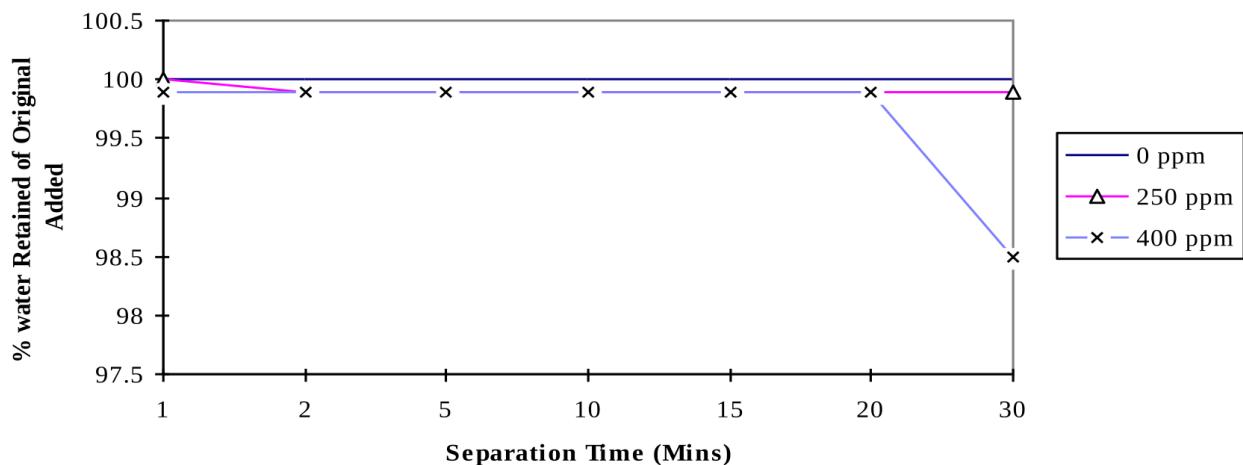
$$A = \pi DL + 0.8\pi D^2 \quad (127)$$

The vessel weight is given by

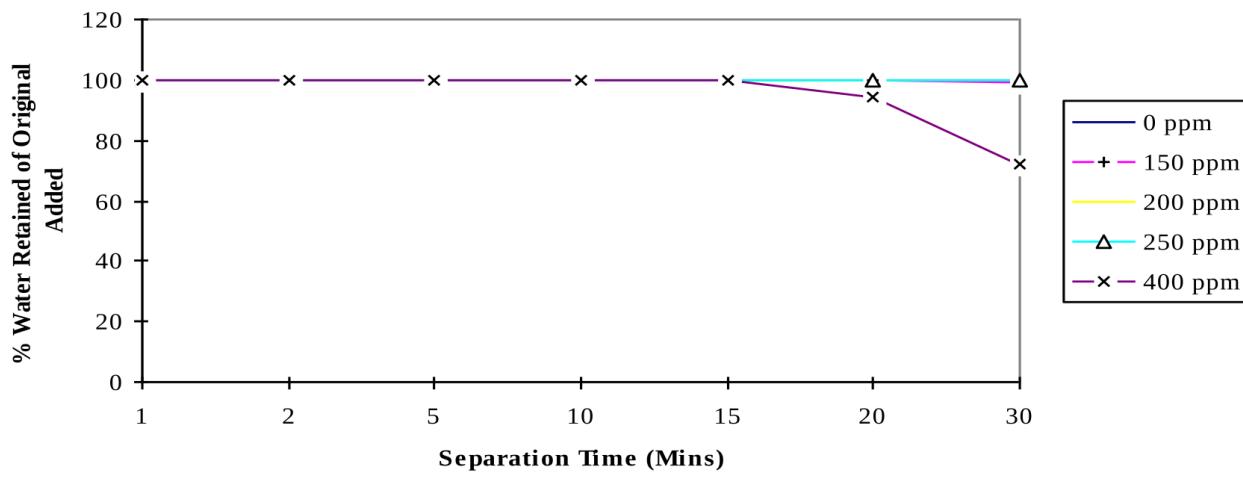
$$W(\text{kg}) = 8000At \quad (128)$$

## Bottle tests

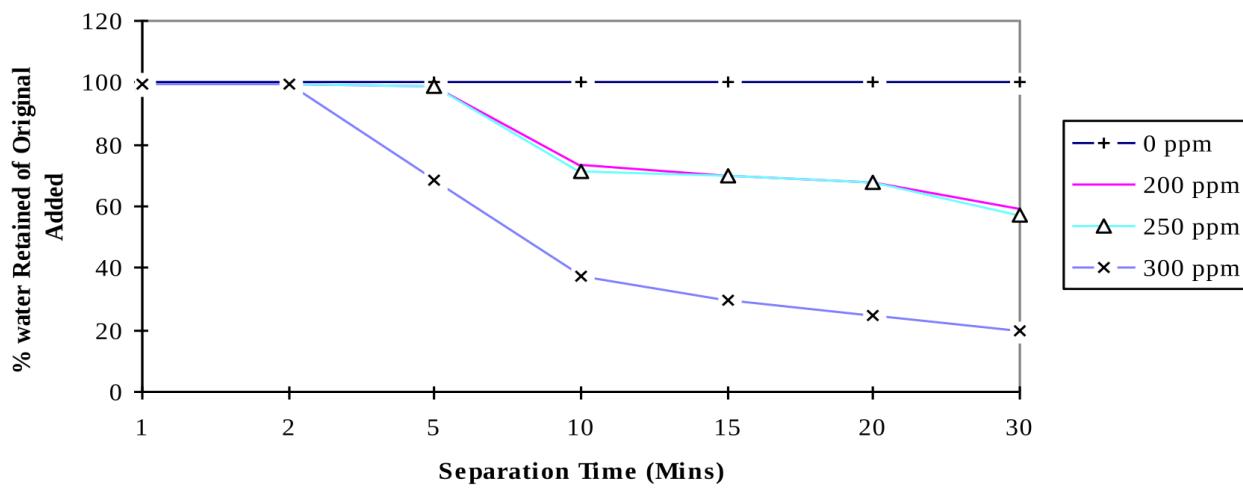
**Clair Crude : 40 Deg. C : TROS 6026a Demulsifier : 10% Water Cut**

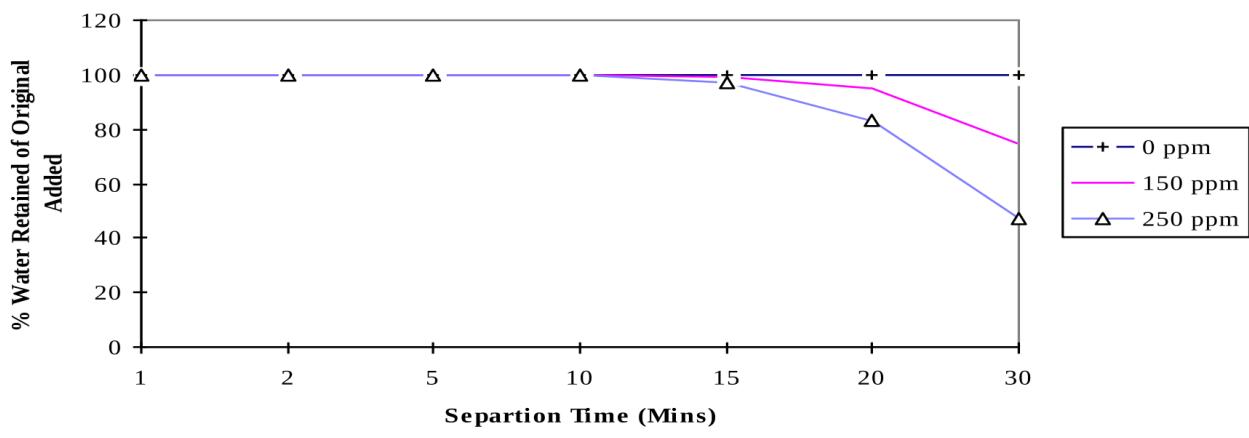
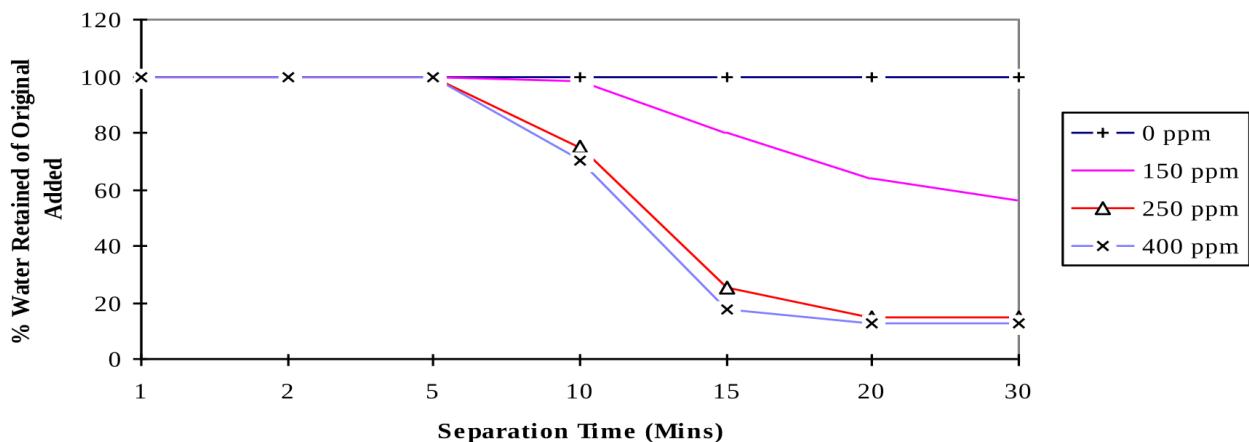
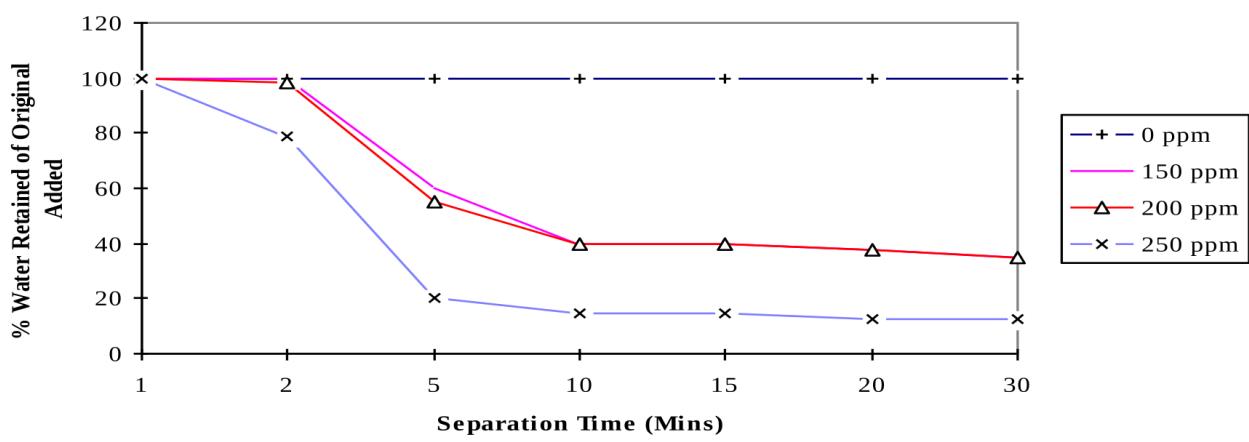


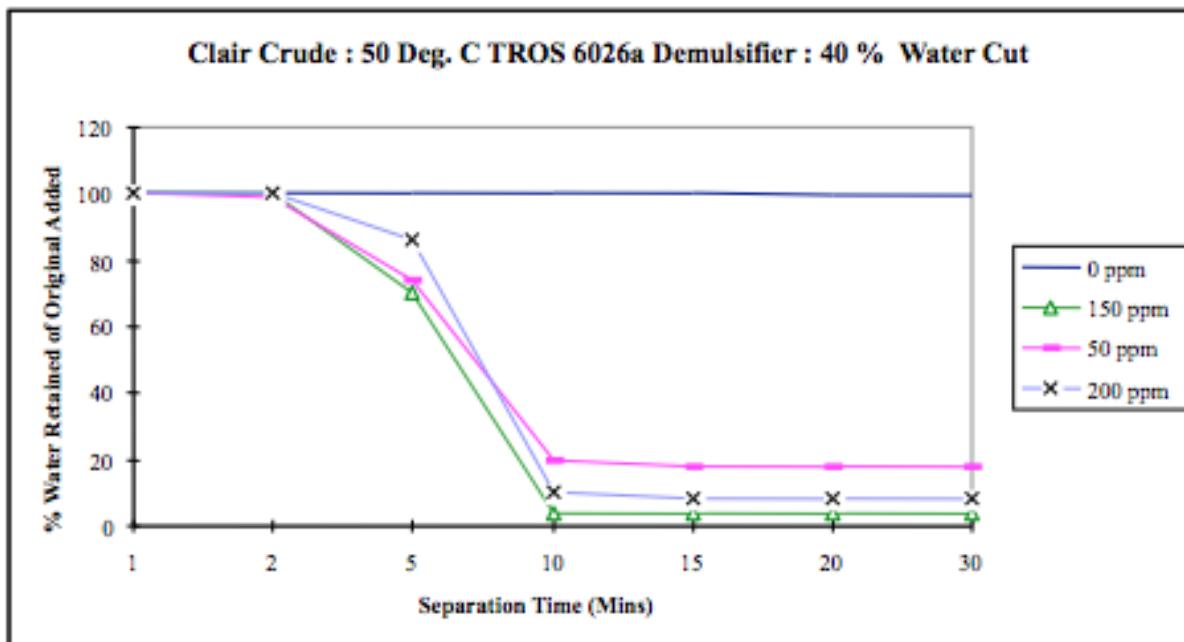
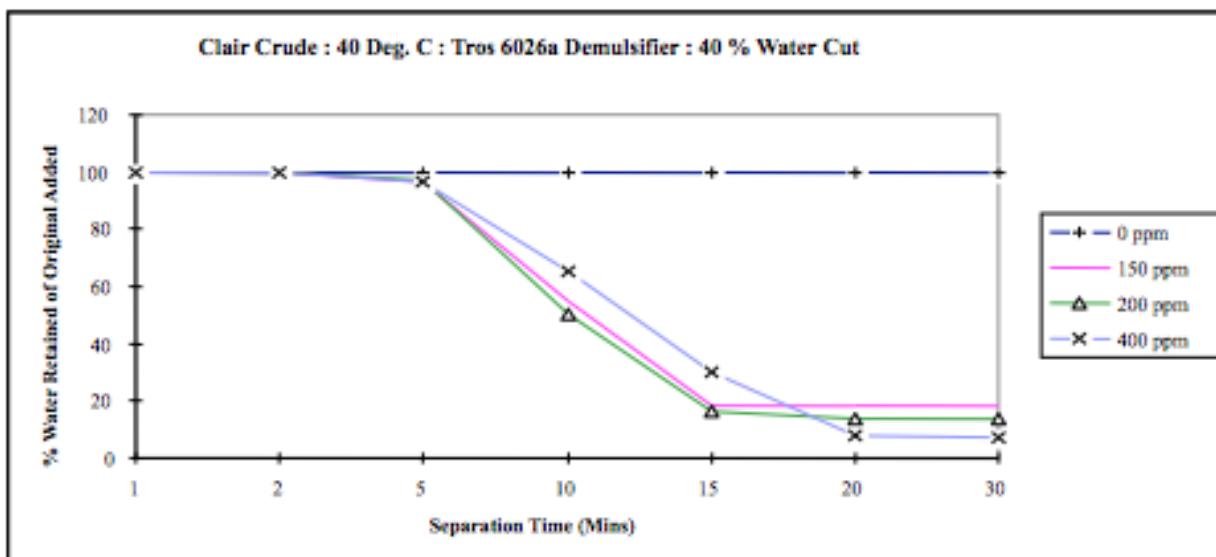
**Clair Crude : 50 Deg. C : TROS 6026a Demulsifier : 10 % Water Cut**



**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

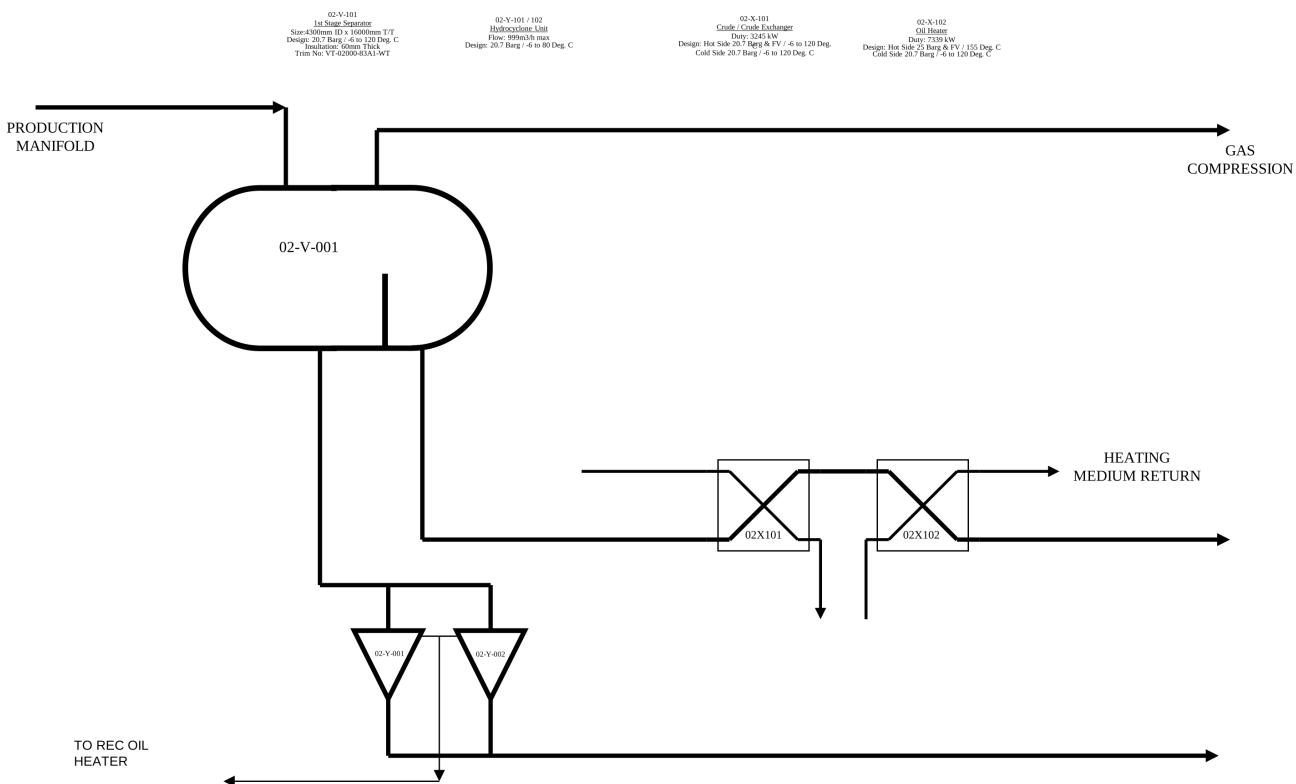
**[Question end]**

## 4 Produced water

**Q.21**

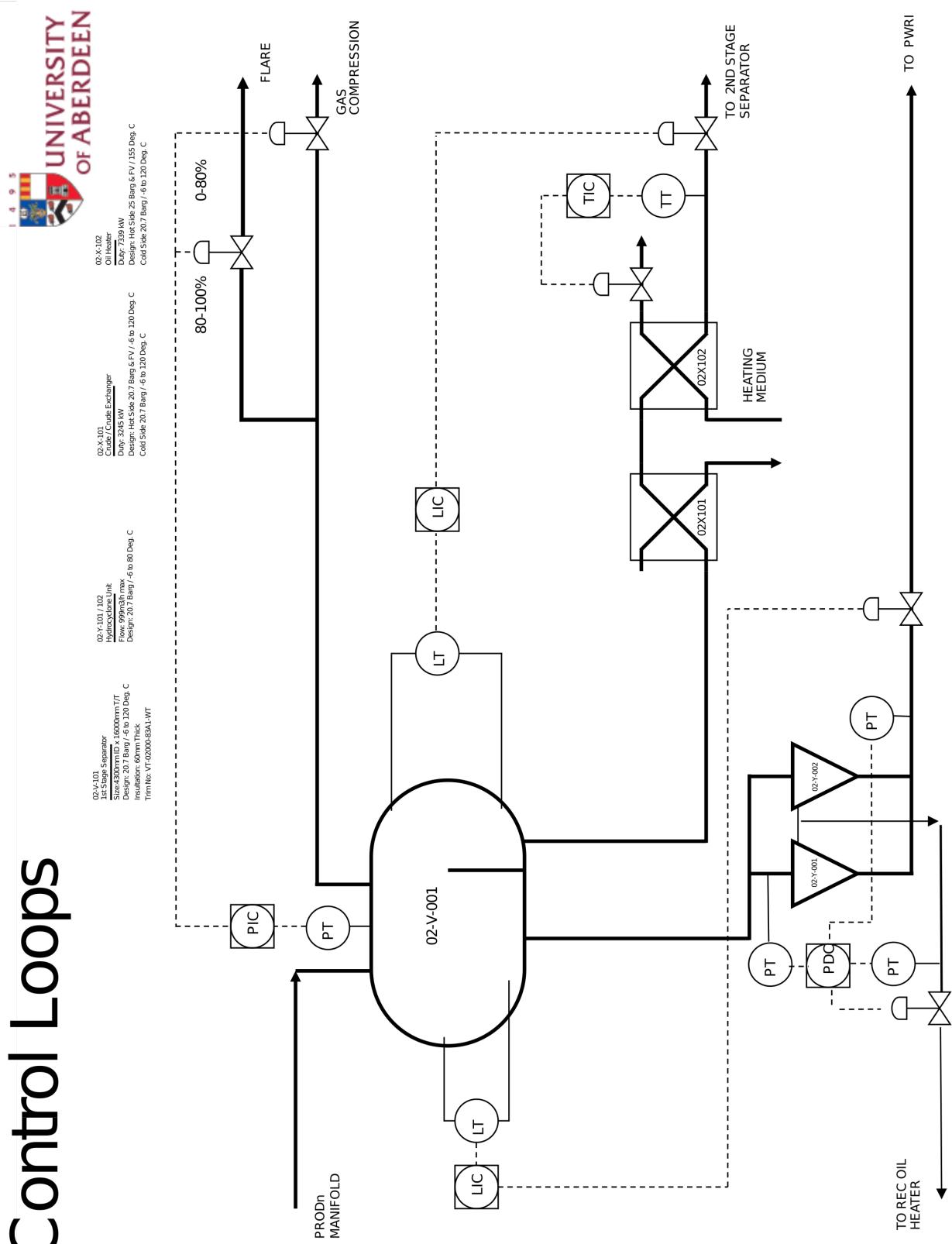
### Question 21

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.



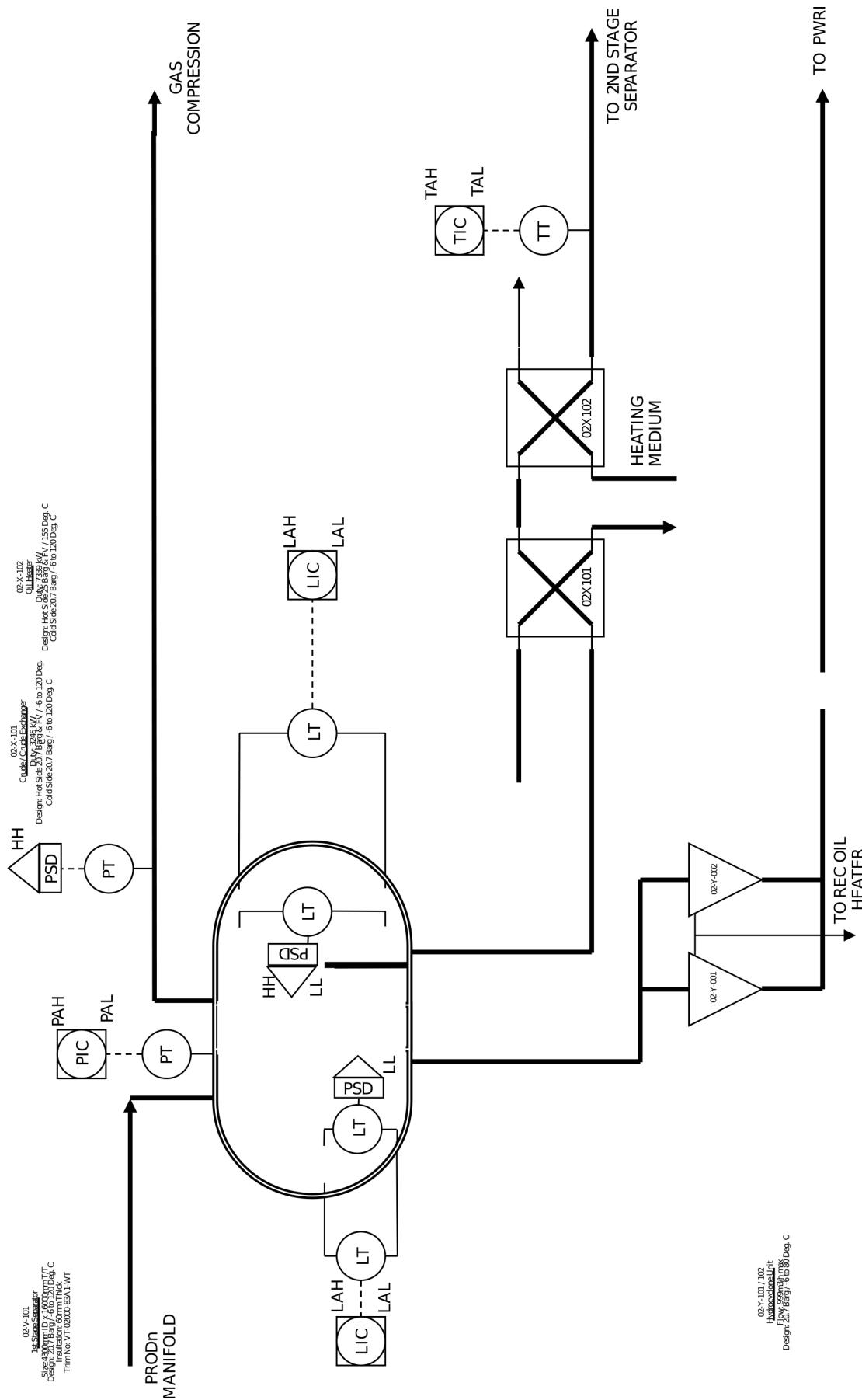
**Solution:**

## Control Loops



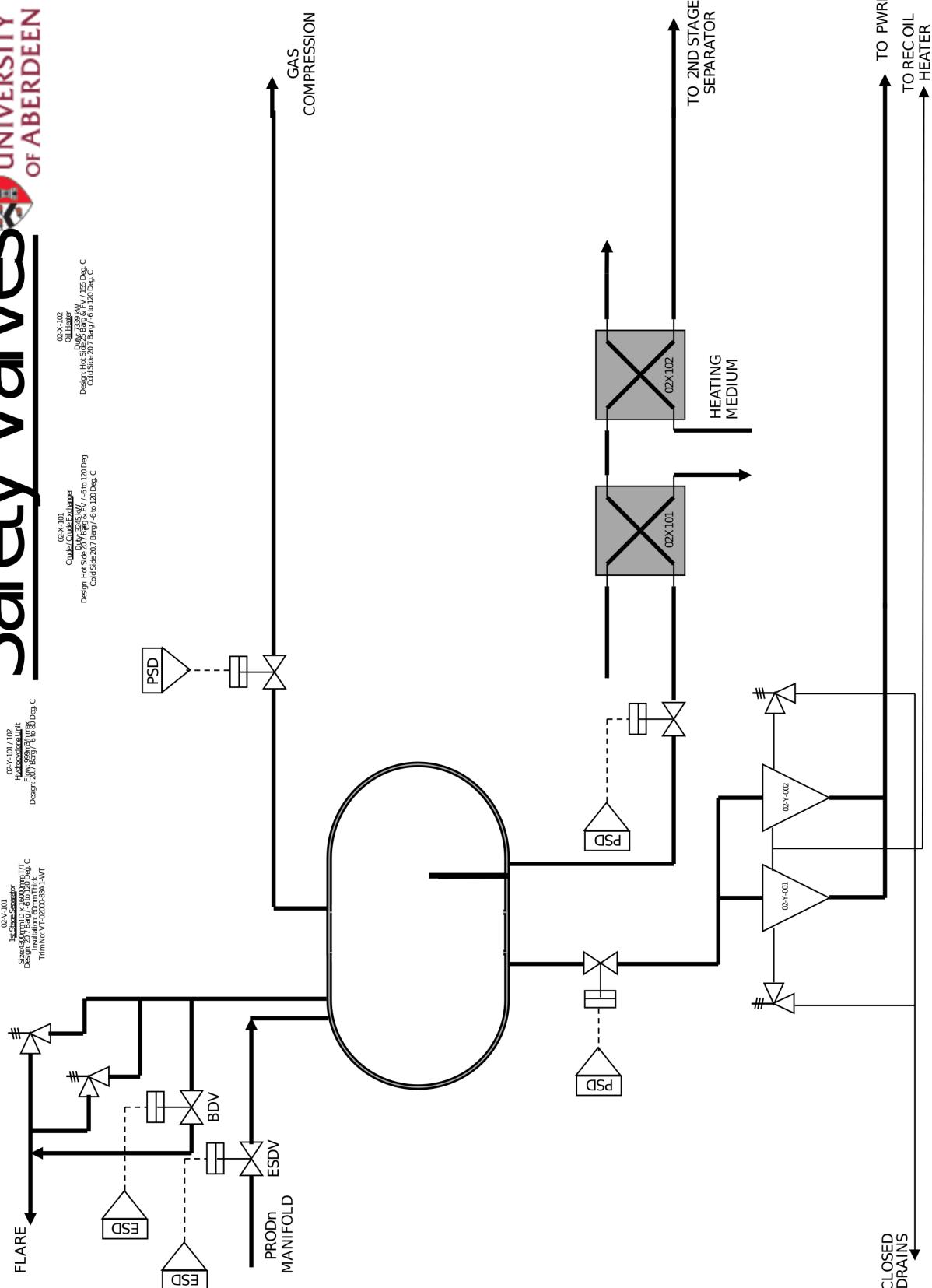


# Alarm & Trip Devices



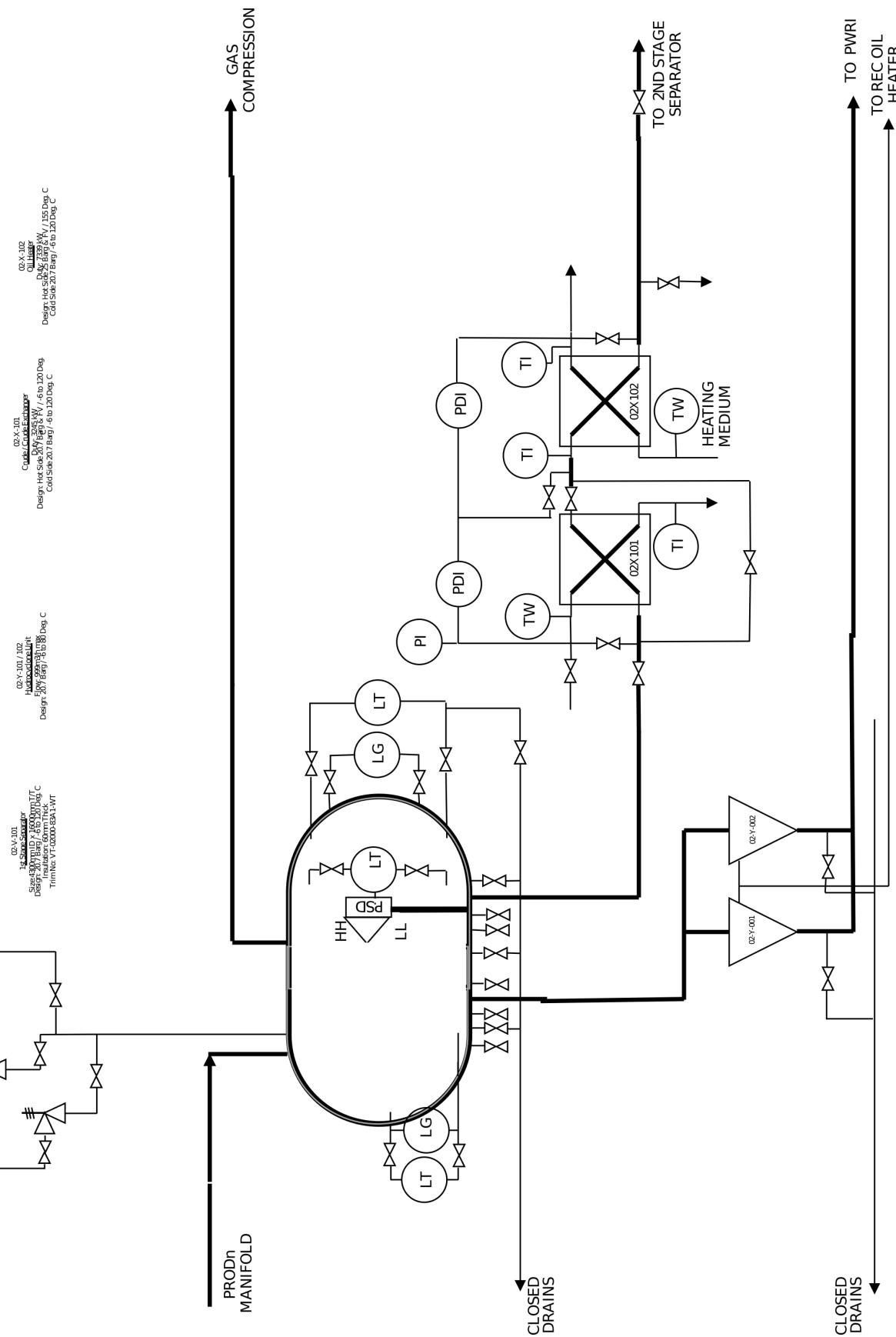


# Safety Valves

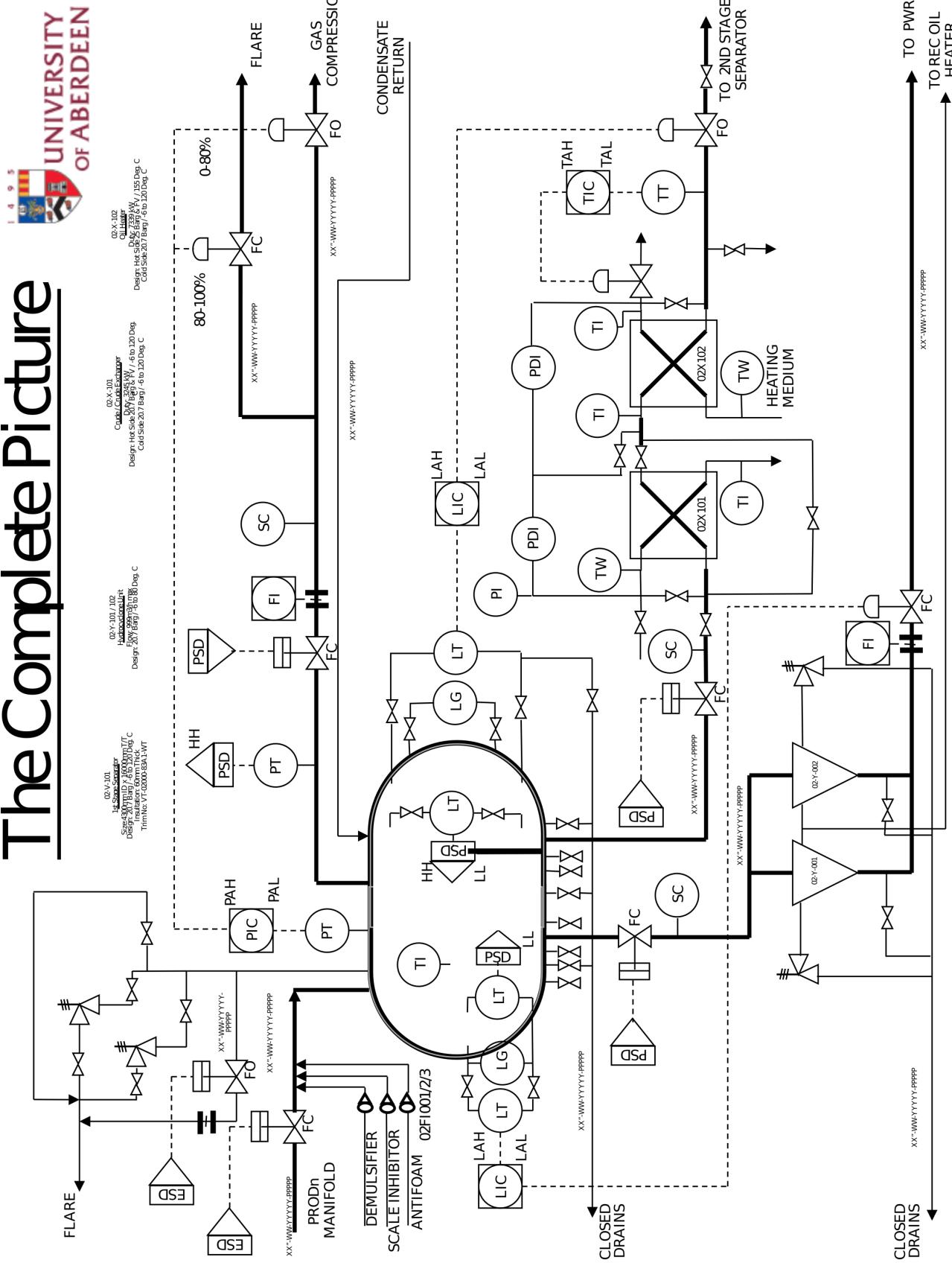


# Maintenance Facilities

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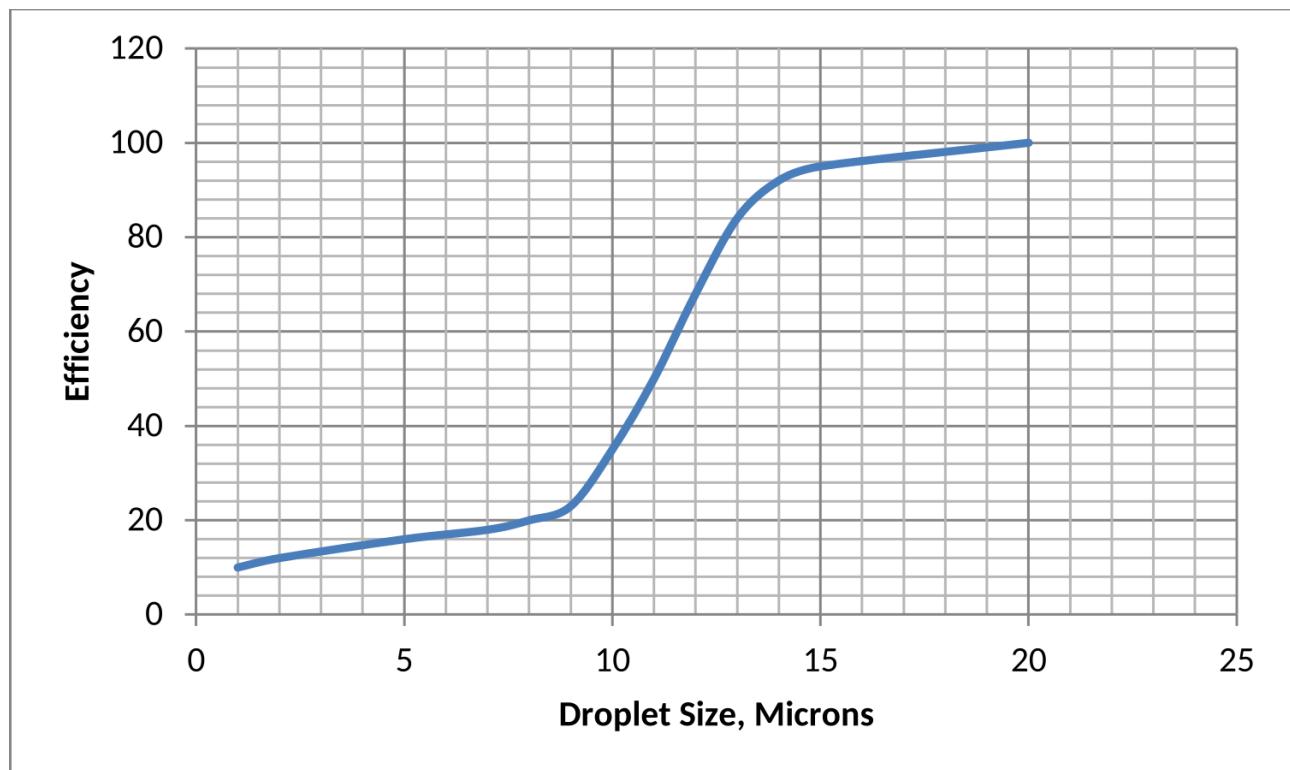
# The Complete Picture



[Question end]

## Q.22 Question 22

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.23 Question 23

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

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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

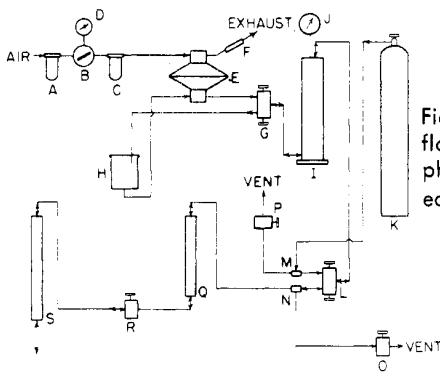


Figure 1. Simplified flow diagram of the phase-contacting equipment

**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  $6 \times 10^{-5}$  ft.<sup>2</sup> per hr; and that of water in benzene,  $3 \times 10^{-4}$ . 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

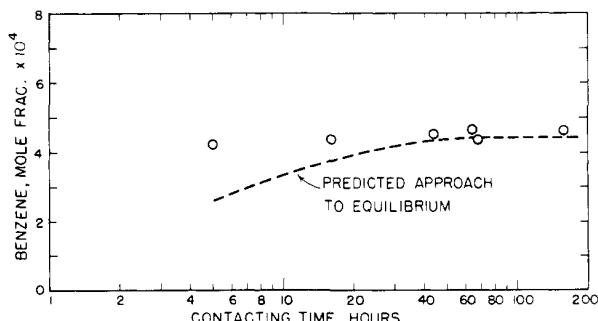


Figure 2. Benzene dissolved in water at 100° F. and 1000 p.s.i.g. as a function of contacting time

Table I. Experimental Solubility and Density Results

Temp., °F.	Benzene Phase		Water Phase	
	Water concn., mole frac. × 10 <sup>4</sup>	Density, g./cc.	Benzene concn., mole frac. × 10 <sup>4</sup>	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

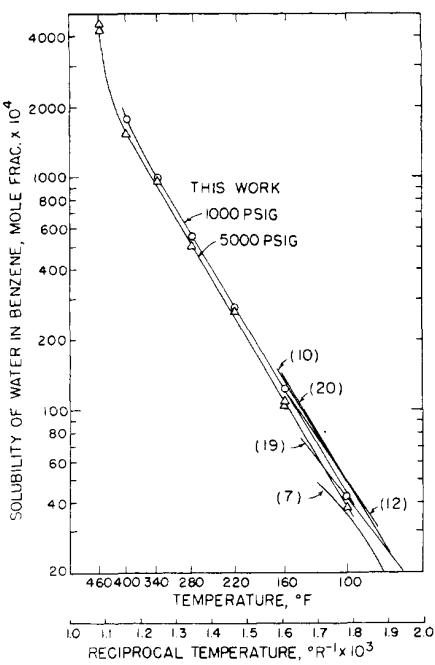
\* = No analysis was performed. † = Discarded data. ‡ = Results known to be high due to a high analytical blank.

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

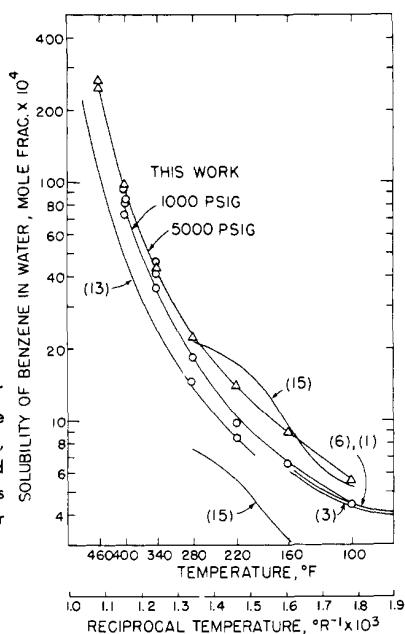
The high-temperature solubility results of Jaeger (13) required conversions and corrections for comparison with the solubilities from this study. In his pioneer work, Jaeger analyzed by simply measuring the volumes which separated after cooling the samples to room temperature. Accordingly room-temperature 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,  $4 \times 10^{-4}$  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.

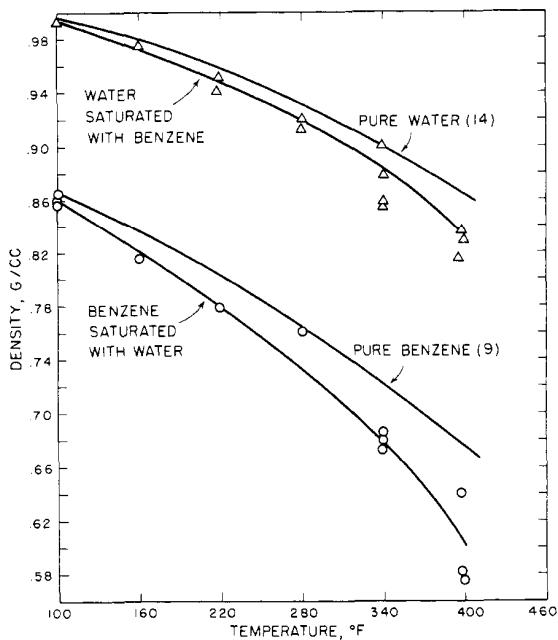
The two isobars that were plotted in Figure 4 from the smoothed results of Kudchadker and McKetta (15) do not agree with this work or the work of others. These two isobars (for 60 and 450 p.s.i.a) indicate very unusual tem-



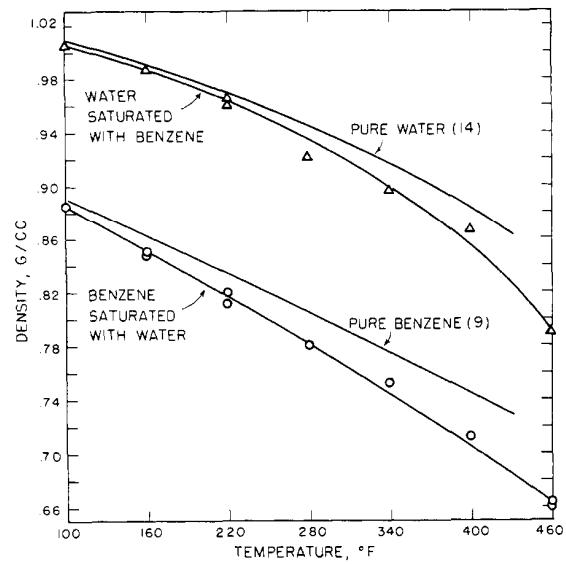
**Figure 3. Solubility of water in benzene**  
The following work of others was performed at one atmosphere pressure or less: Clifford (7), Groschuff (10), Hill (12), Rosenbaum and Walton (19), and Stavely, Jeffes, and Moy (20).



**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		
	Water concn.			Benzene concn.		
	Mole frac. × 10 <sup>4</sup>	Wt. frac. × 10 <sup>4</sup>	Moles/liter	Density, g./cc.	Mole frac. × 10 <sup>4</sup>	Wt. frac. × 10 <sup>4</sup>
1000 P.S.I.G.						
100	42.3	9.78	0.0467	0.860	4.46	19.3
160	121	28.2	0.129	0.821	6.65	28.8
220	274	64.5	0.279	0.779	10.1	43.6
280	558	134	0.544	0.732	18.6	80.1
340	1000	250	0.944	0.679	36.0	154
400	1780	475	1.58	0.601	82.0	346
5000 P.S.I.G.						
100	38.1	8.80	0.0432	0.885	5.61	24.3
160	106	24.6	0.116	0.851	9.05	39.2
220	245	57.5	0.261	0.817	14.1	60.8
280	500	120	0.520	0.781	22.8	98.0
340	905	224	0.927	0.745	43.0	184
400	1620	445	1.74	0.706	98.0	411
460	4500	1590	5.83	0.663	255	1020

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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## 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

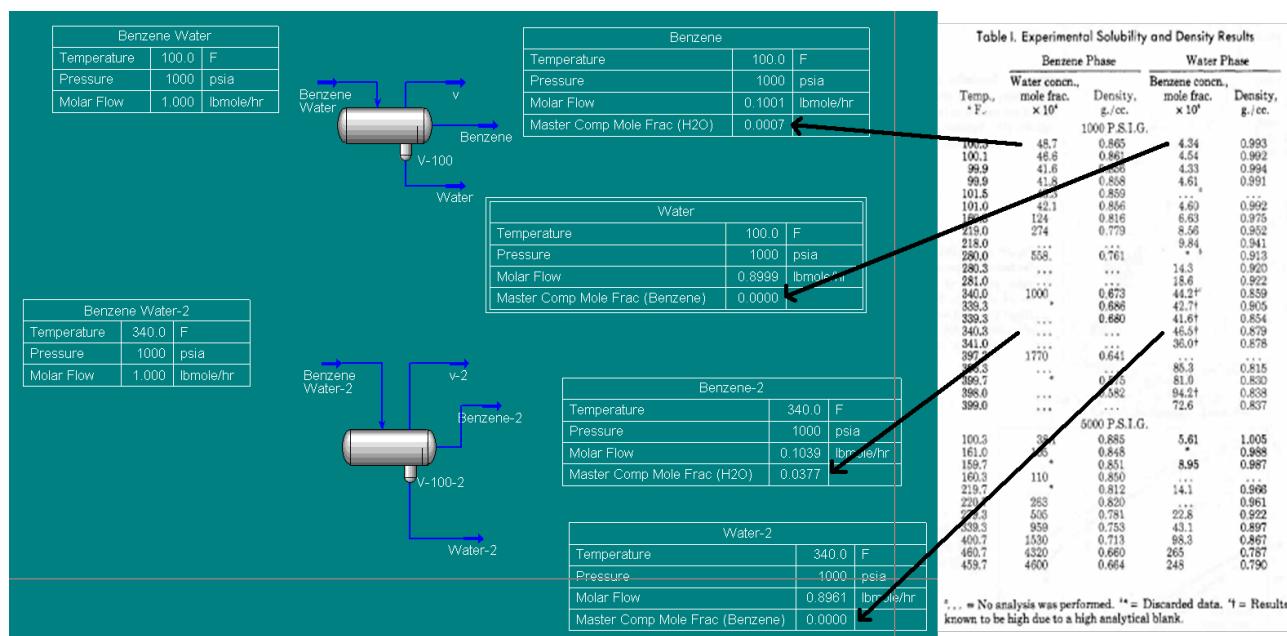
**Rubidium Samples.** Two batches of pure rubidium were obtained commercially for use in this program. Cesium represented the major impurity in each batch, being present to the extent of 0.4 and 0.02 wt. per cent, respectively. The rubidium was analyzed for trace metal constituents by emission spectroscopy;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

## Solution:

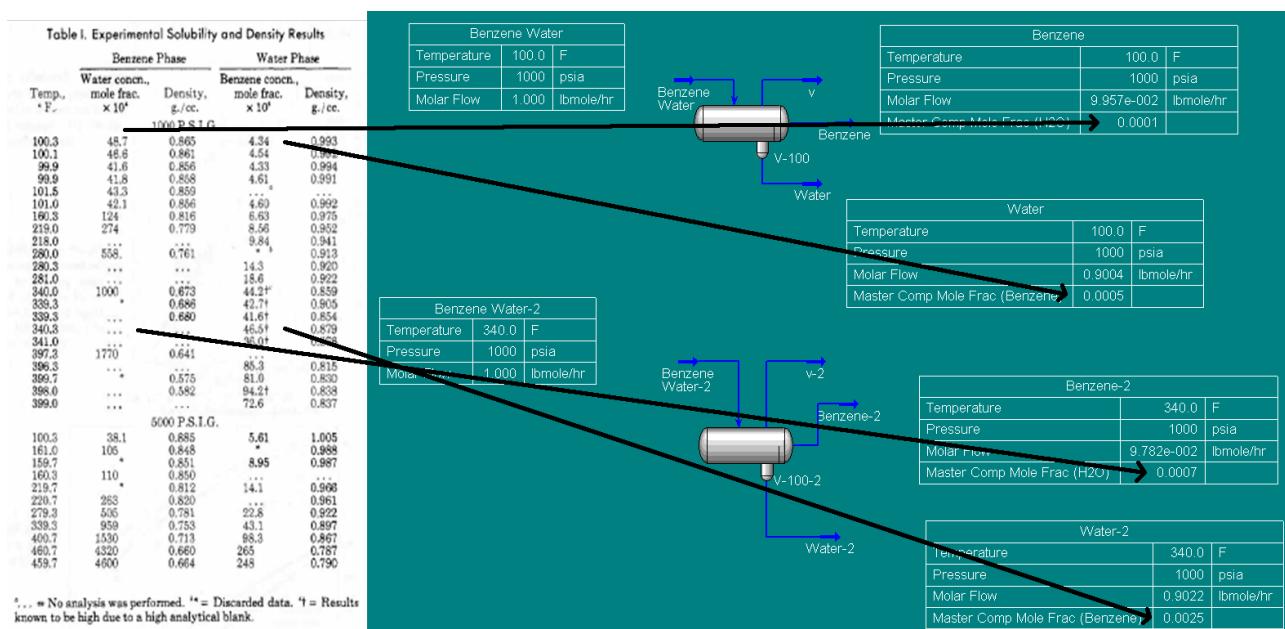
Table. 1 in the attached paper provides data which can be compared with phase behaviour models. A comparison is made on UniSim using Peng Robinson and NRTL (non random two liquid). Benzene and water are combined in a 3 phase separator where the water phase will saturate with benzene and vice versa.

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

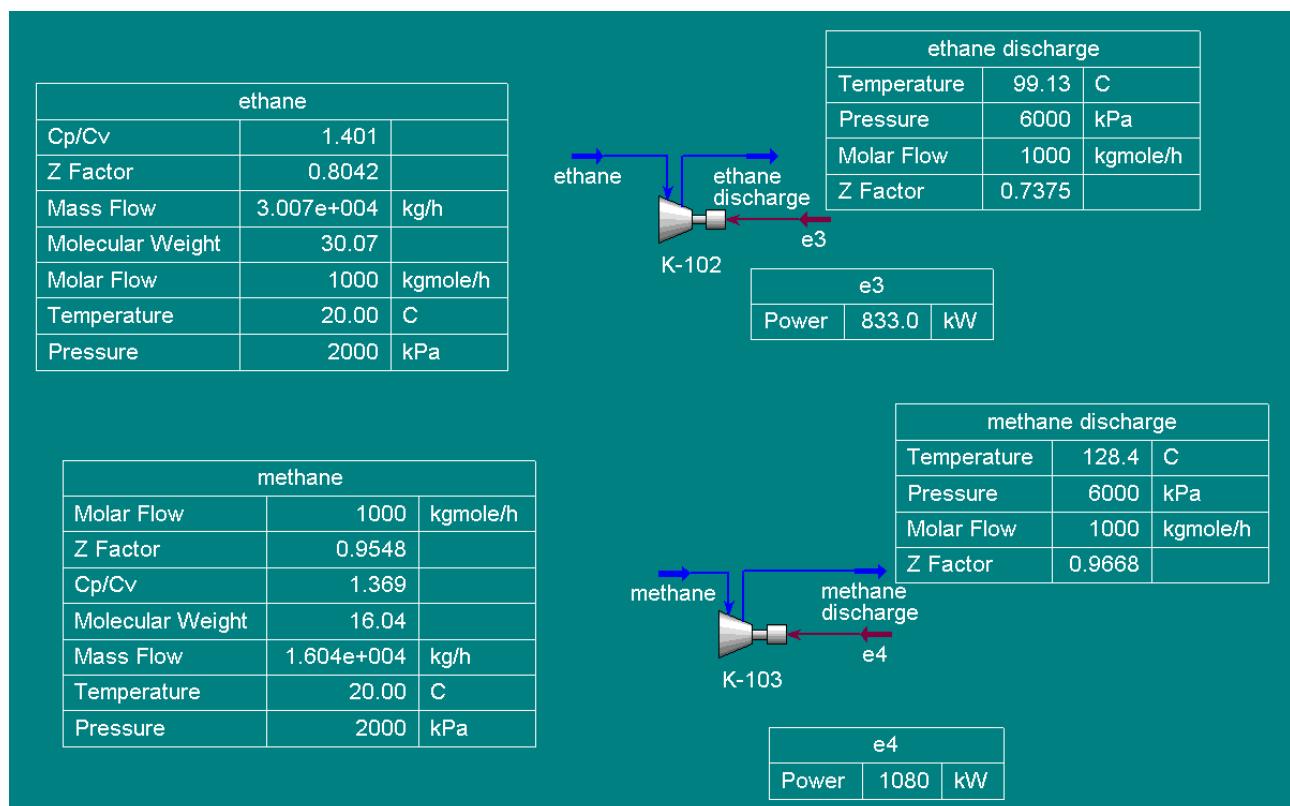


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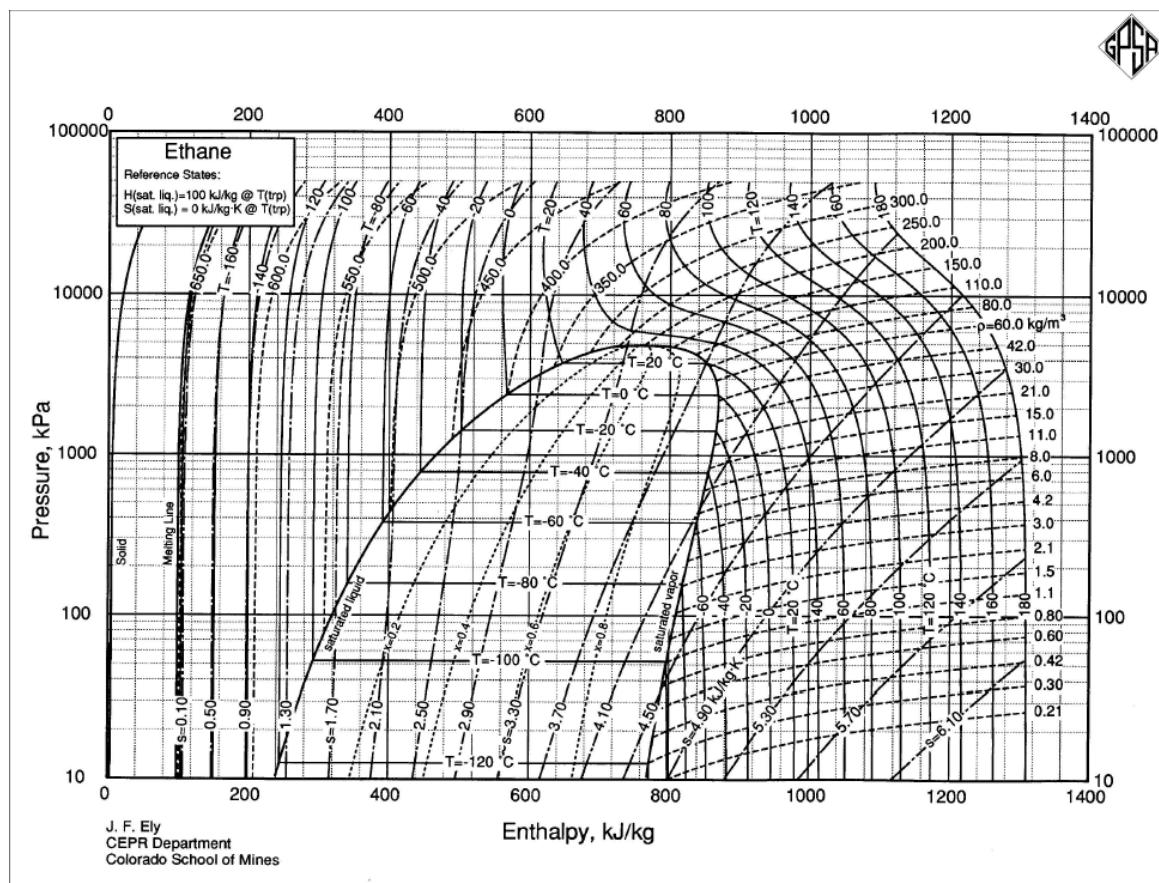
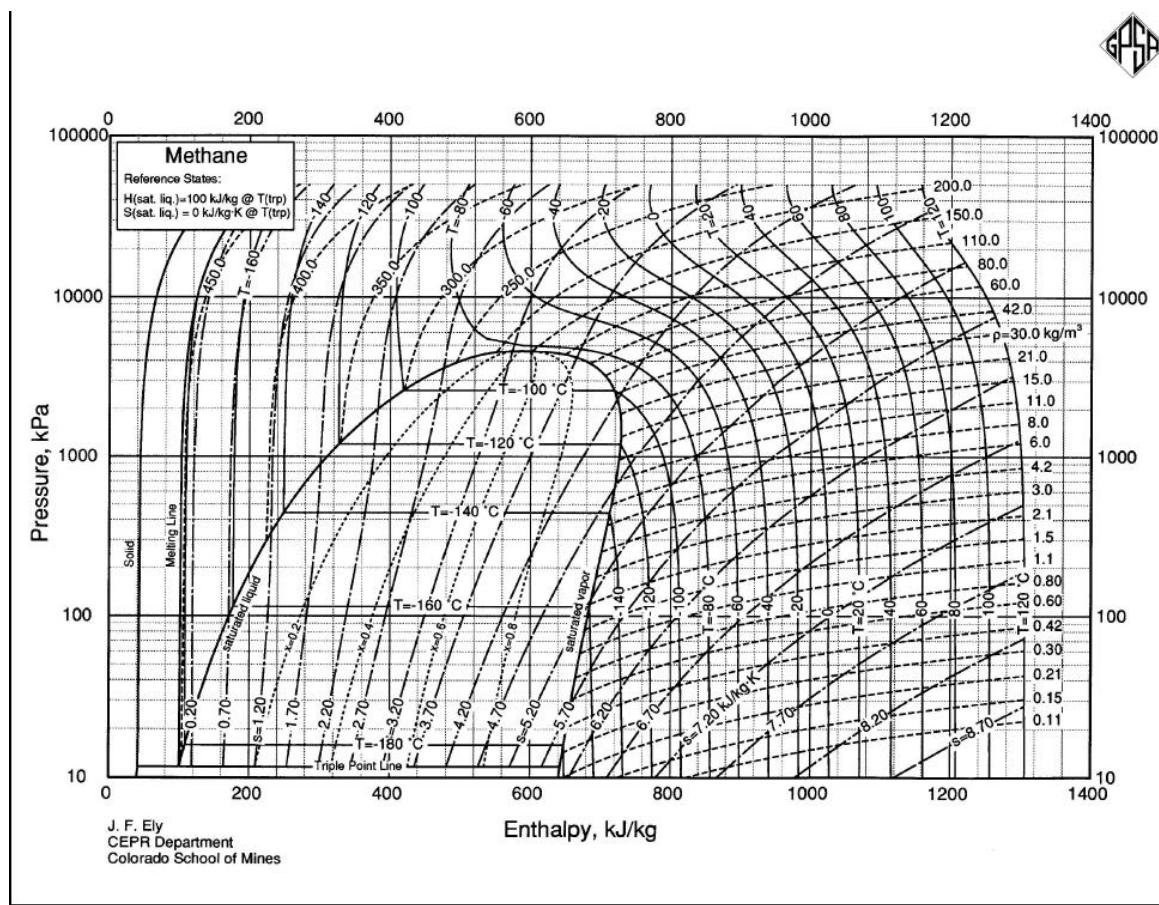
## 5 Gas compression

### Q.24 Question 24

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:

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

$\Delta h_{poly}$	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)
$\phi_m$	Mass flow (kg/s)
$E_{poly}$	Polytropic efficiency - Function of volumetric inlet flow varying from approx. 0.6 to 0.8 (compressor specific)
$\gamma$	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	Pressure (psia)	Critical Constants 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 <sup>a</sup>	0.969	742	50
Propane	44.09	1.13	0.58	1.522	617	206
Propylene	47.08	1.15	0.58 <sup>a</sup>	1.453	667	197
Isobutane	58.12	1.18	0.59 <sup>a</sup>	2.007	529	273
n-Butane	58.12	1.19	0.59	2.007	551	304
1-Butene	56.10	1.11	0.59 <sup>a</sup>	1.937	543	276
Isopentane	72.15	1.08	0.59 <sup>a</sup>	2.491	483	369
n-Pentane	72.15	1.08	0.59 <sup>a</sup>	2.491	490	386
1-Pentene	70.13	1.08	0.59 <sup>a</sup>	2.421	586	377
n-Hexane	86.18	1.06	0.59 <sup>a</sup>	2.973	437	454
Benzene	78.11	1.12	0.58	2.697	714	552
n-Heptane	100.20	1.05	0.60 <sup>a</sup>	3.459	397	513
Toluene	92.13	1.09	0.59	3.181	590	604
n-Octane	114.22	1.05	0.60 <sup>a</sup>	3.944	362	564
n-Nonane	128.23	1.04	0.60 <sup>a</sup>	4.428	552	610
n-Decane	142.28	1.03	0.60 <sup>a</sup>	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

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

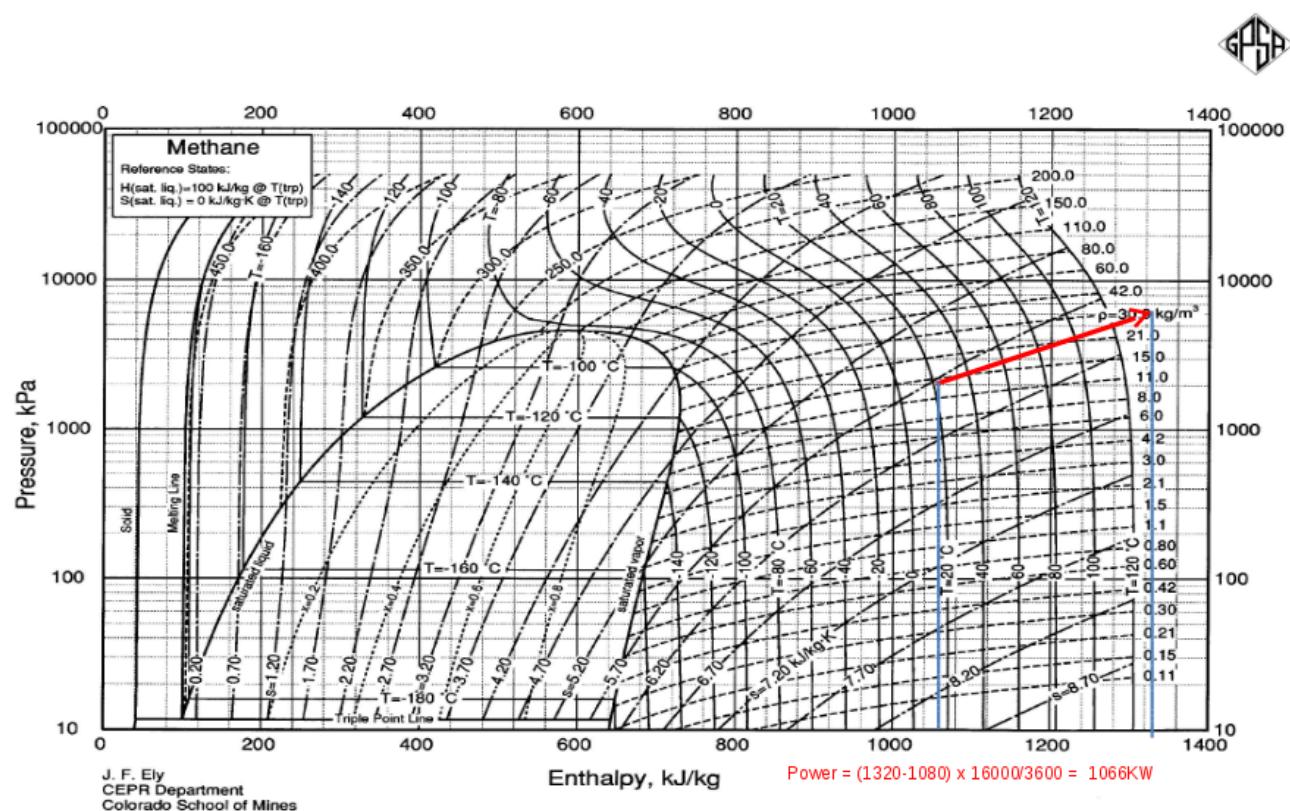
This equation is rearranged for the compressor efficiency  $E$  and then used to solve for the  $E$  which gives an output temperature of  $T = 160 \text{ }^{\circ}\text{C} = 433 \text{ K}$  given that  $(n-1)/n \approx (k-1)/kE$

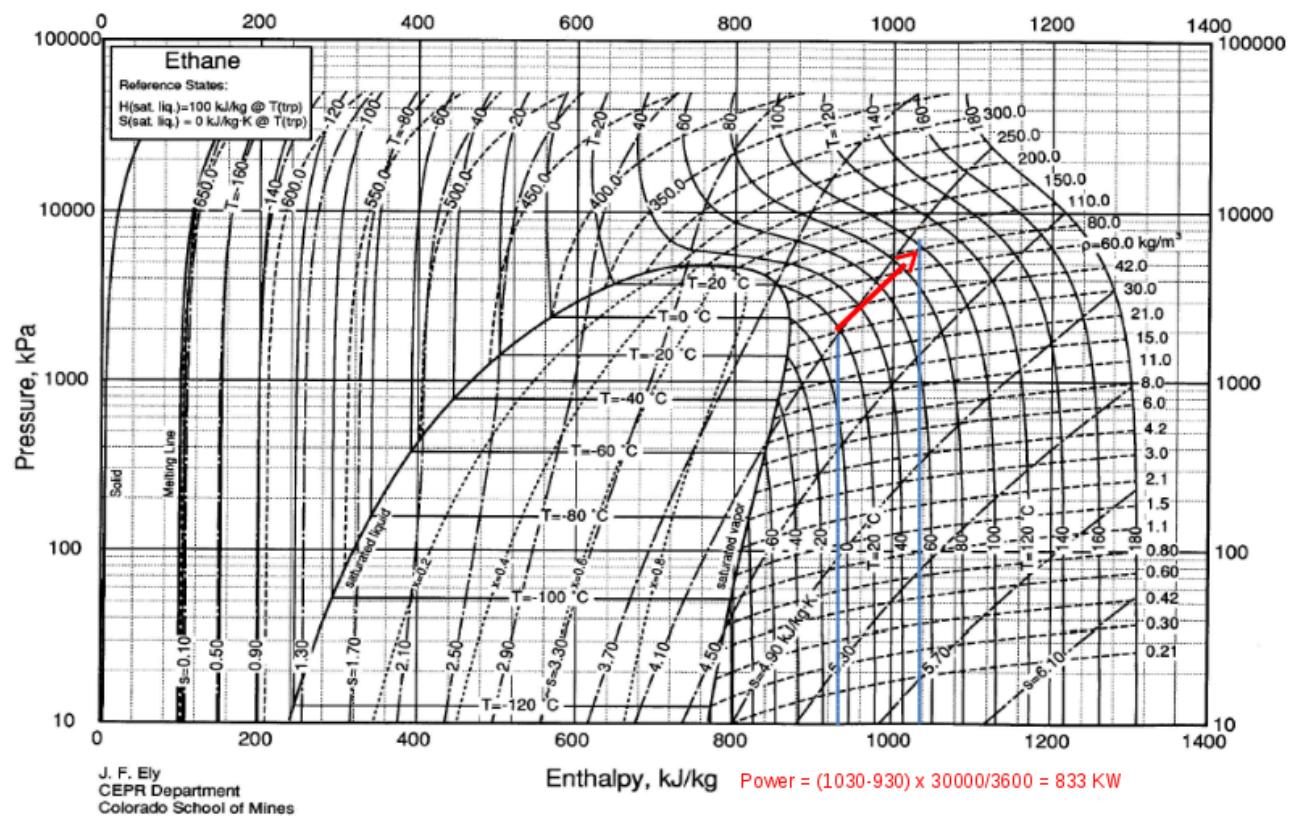
$$E = \frac{k-1}{k} \log_{\frac{T_2}{T_1}} \left( \frac{p_2}{p_1} \right) \quad (150)$$

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

$$= 0.666 \quad (152)$$

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

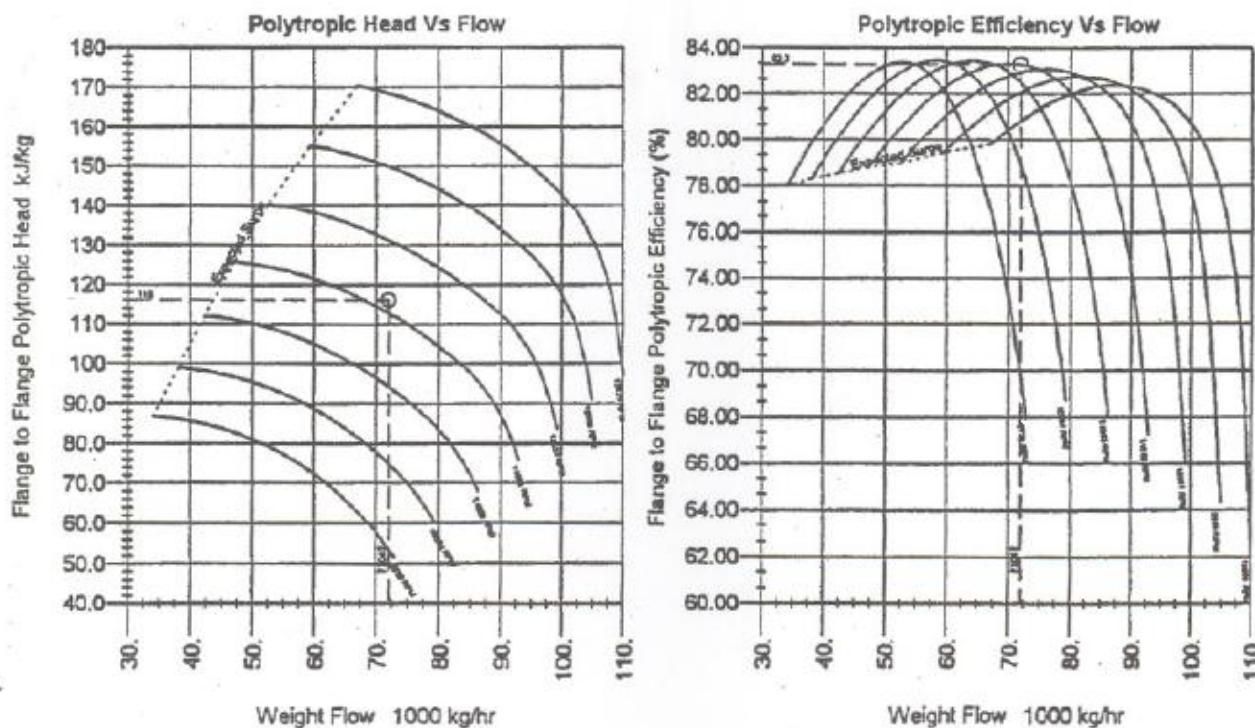




[Question end]

### Q.25 Question 25

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.

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

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

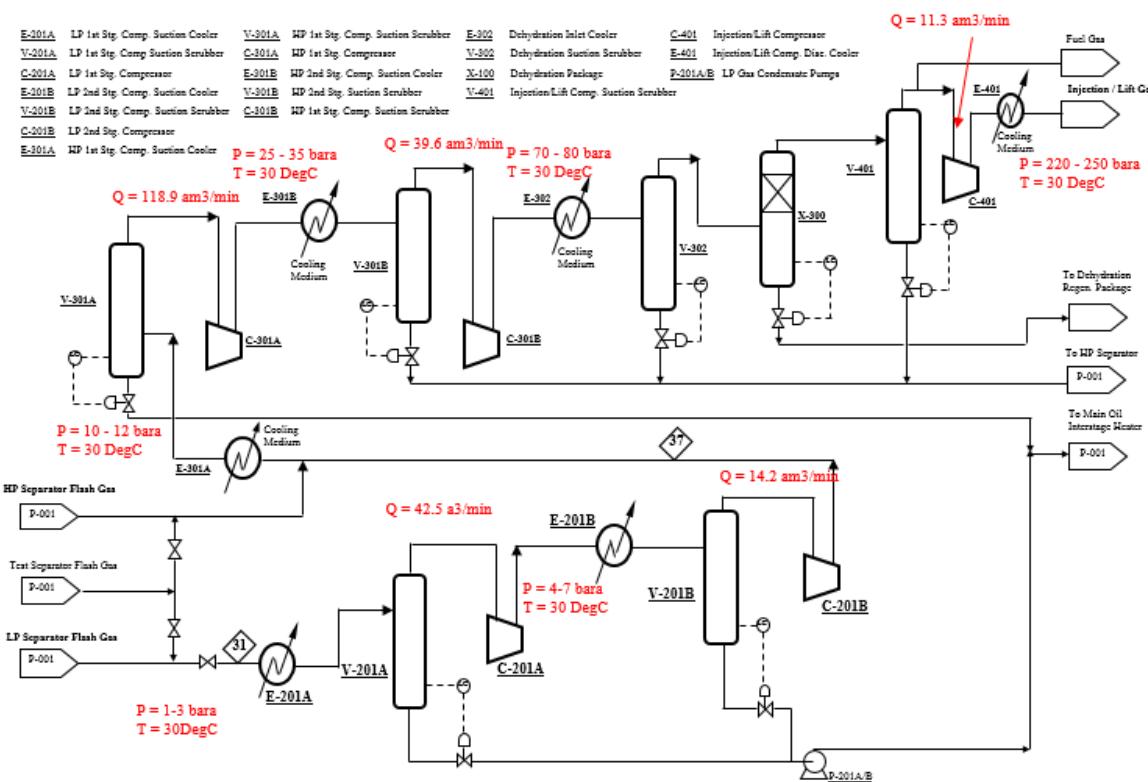
$$= 288 \text{ K} = 15 \text{ }^\circ\text{C} \quad (160)$$

[Question end]

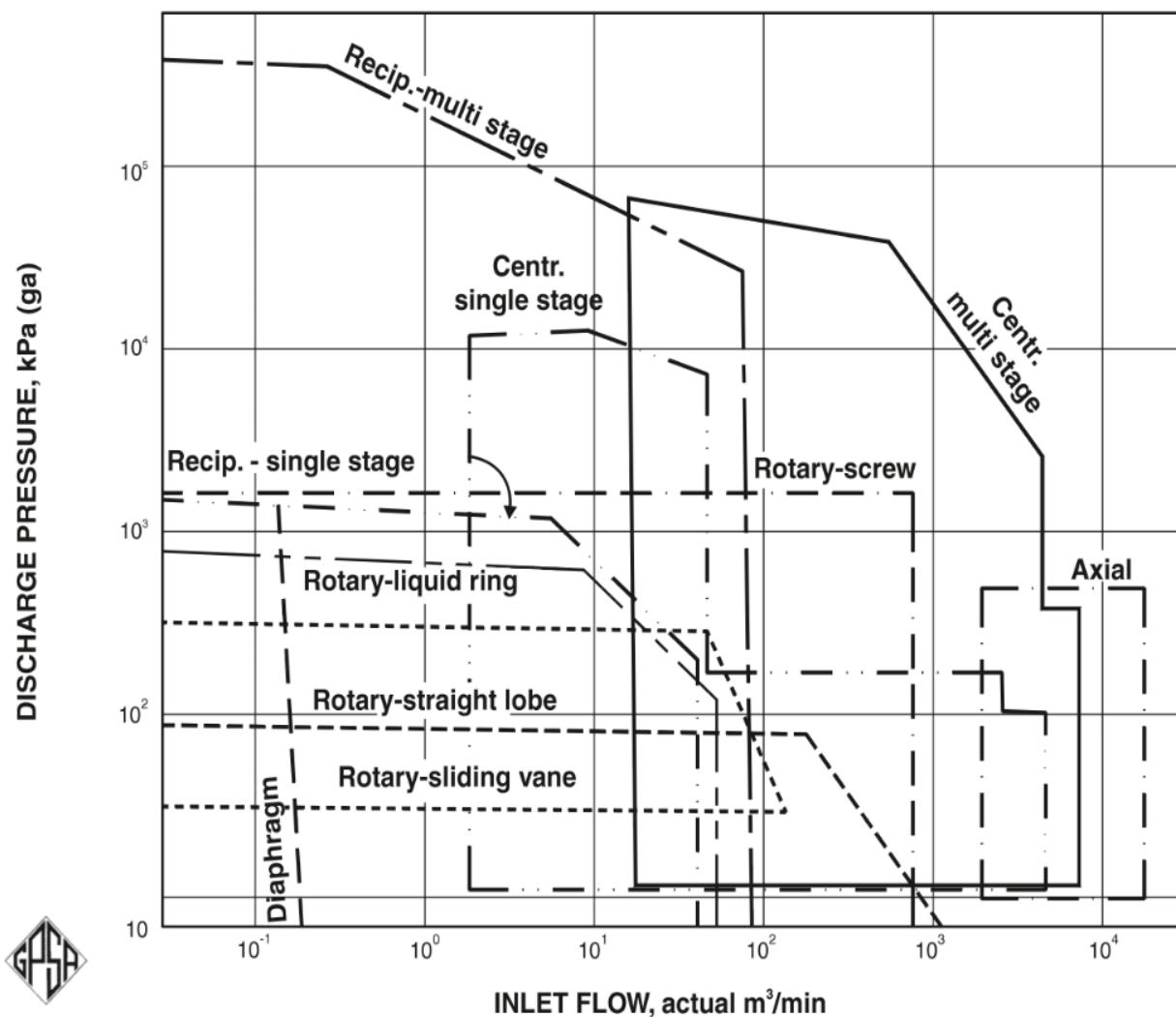
Q.26

**Question 26**

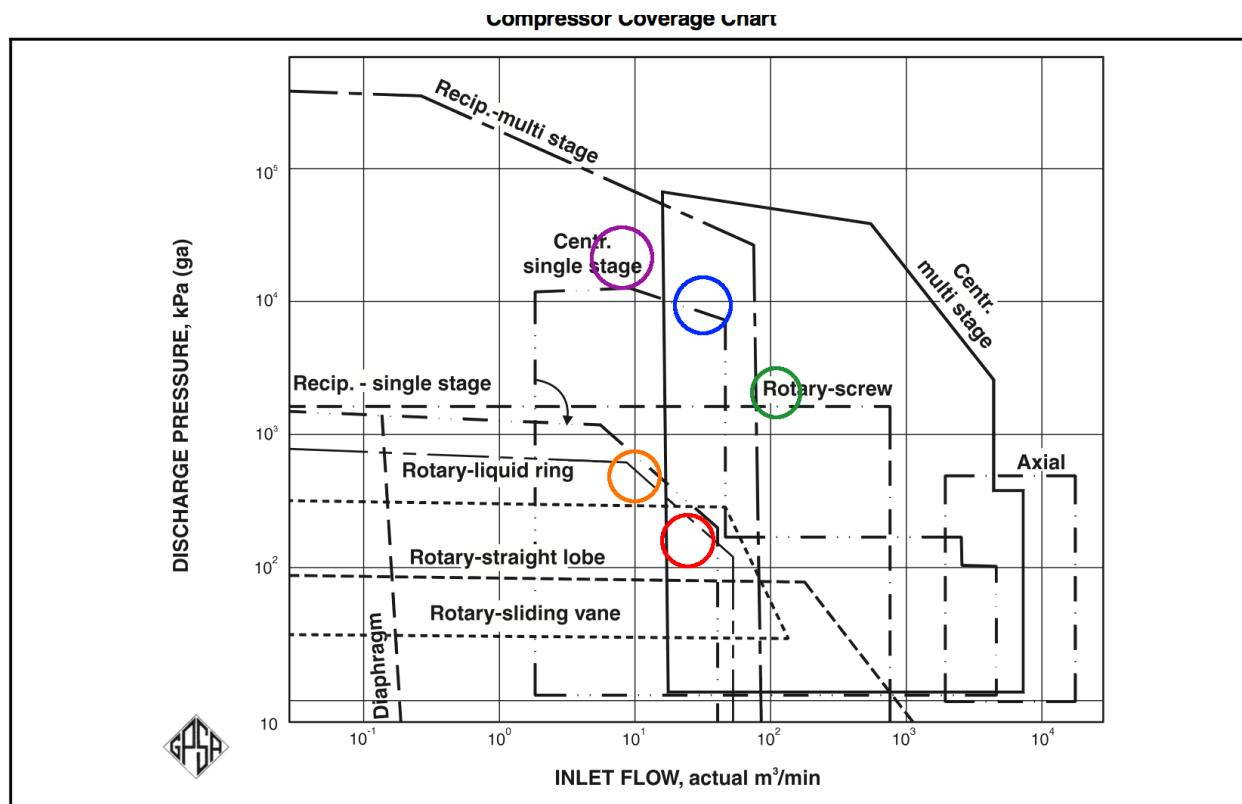
The attached PFD shows a compression train.

**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.

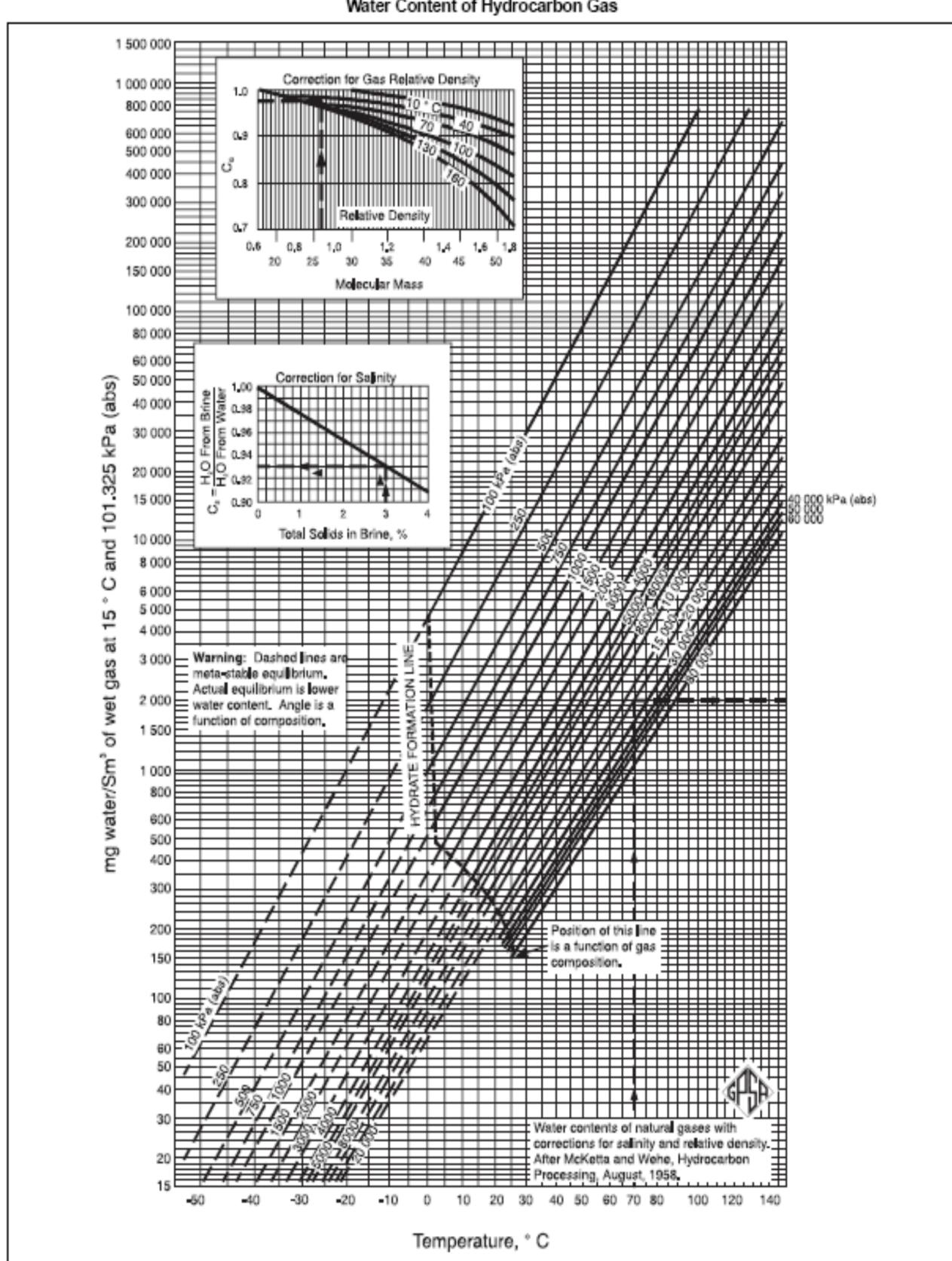
[Question end]

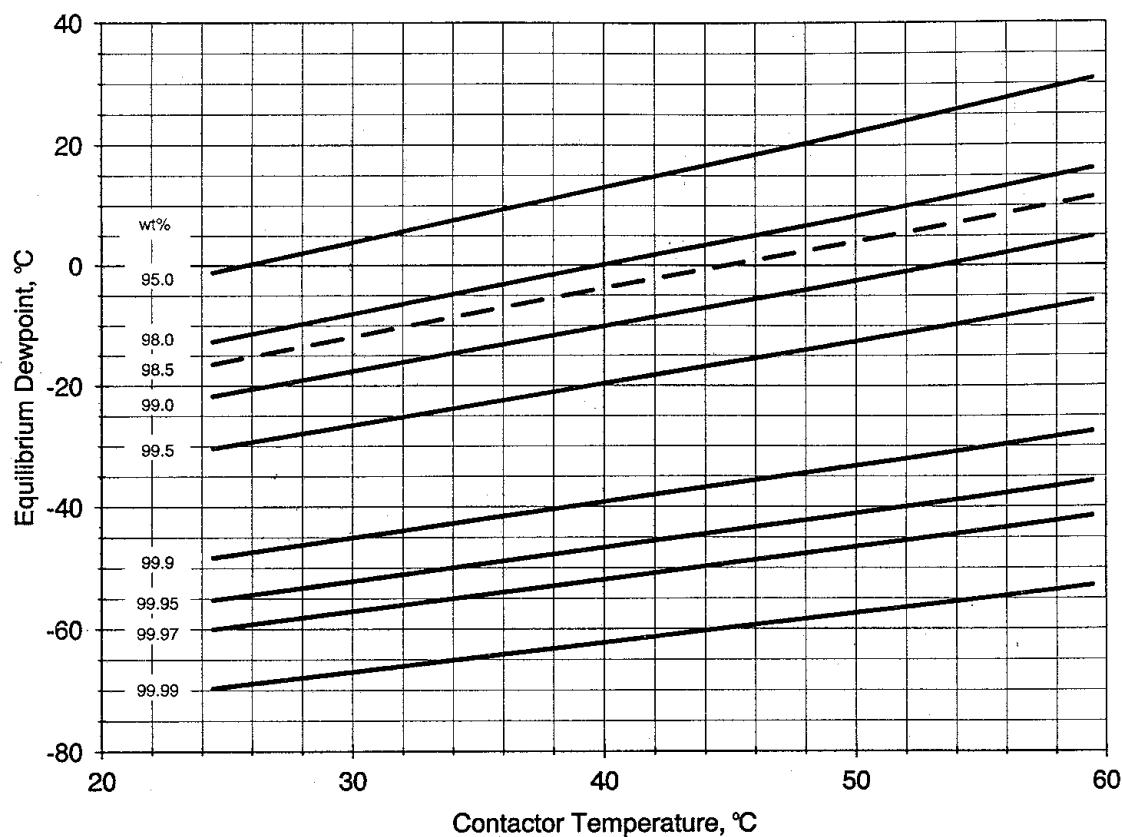
## 6 Gas treatment

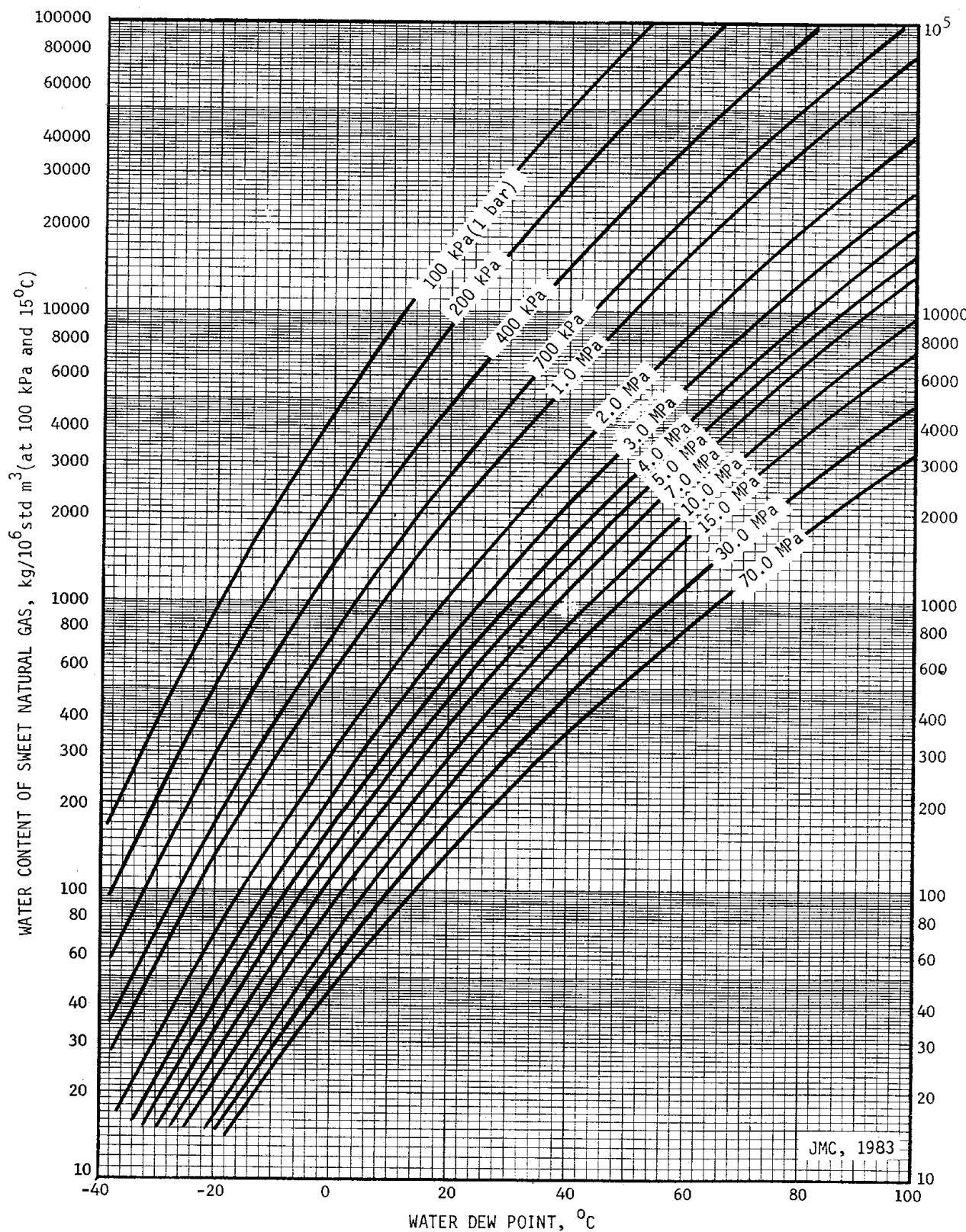
### Q.27 Question 27

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 H<sub>2</sub>O sm<sup>-3</sup>. The flowrate of the gas is  $5 \times 10^6$  sm<sup>3</sup> day<sup>-1</sup>. 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<sup>-3</sup>.







### 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<sup>-3</sup>.
2. The outlet gas specification is 100 mg sm<sup>-3</sup>. 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<sup>-3</sup>.
5. Calculate how much water is transferred to the glycol.

$$Q_{H_2O,transferred} = Q_{gas} \times (c_{H_2O,in} - c_{H_2O,out}) \quad (161)$$

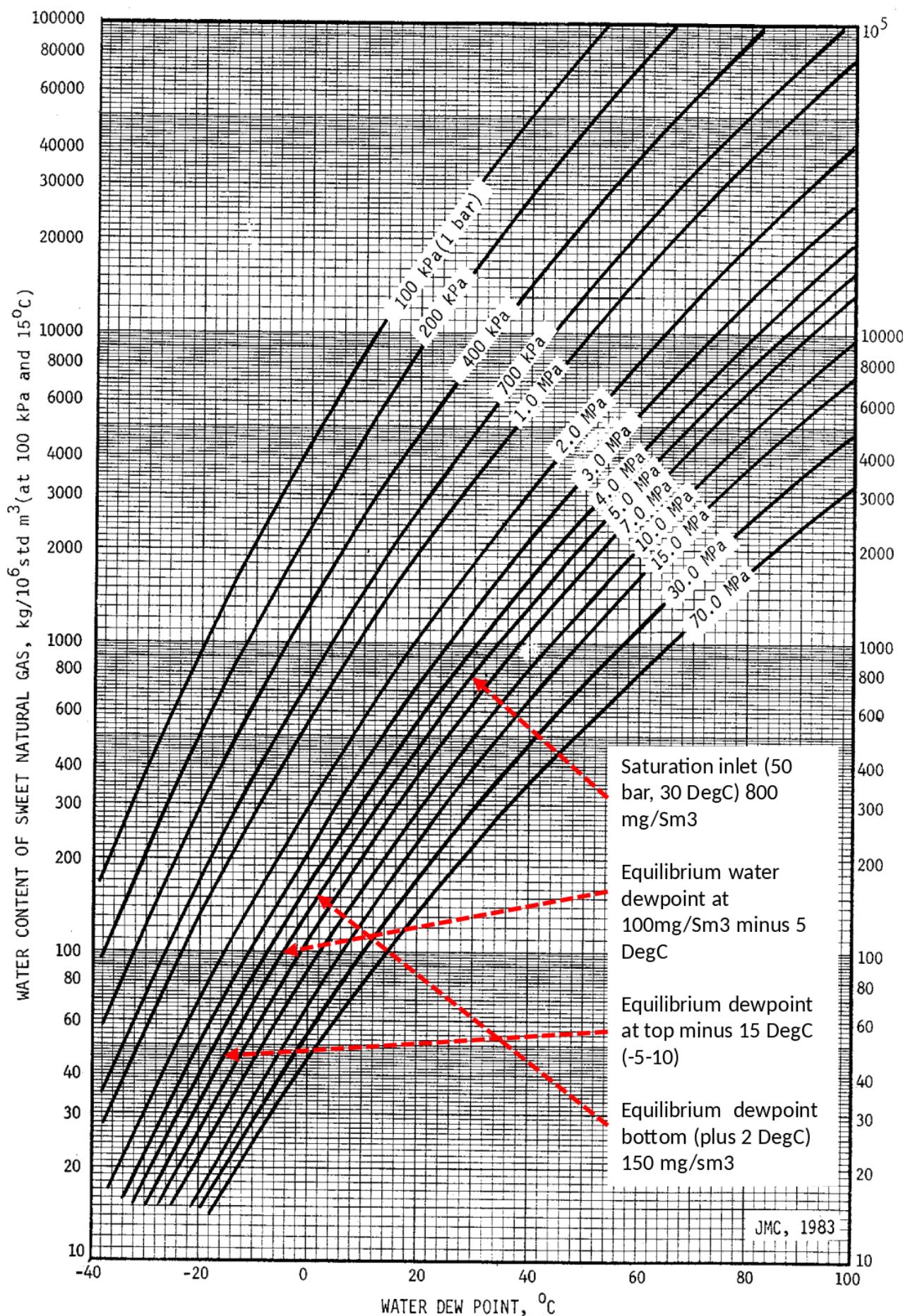
$$= 0.0405 \text{ kg s}^{-1} \quad (162)$$

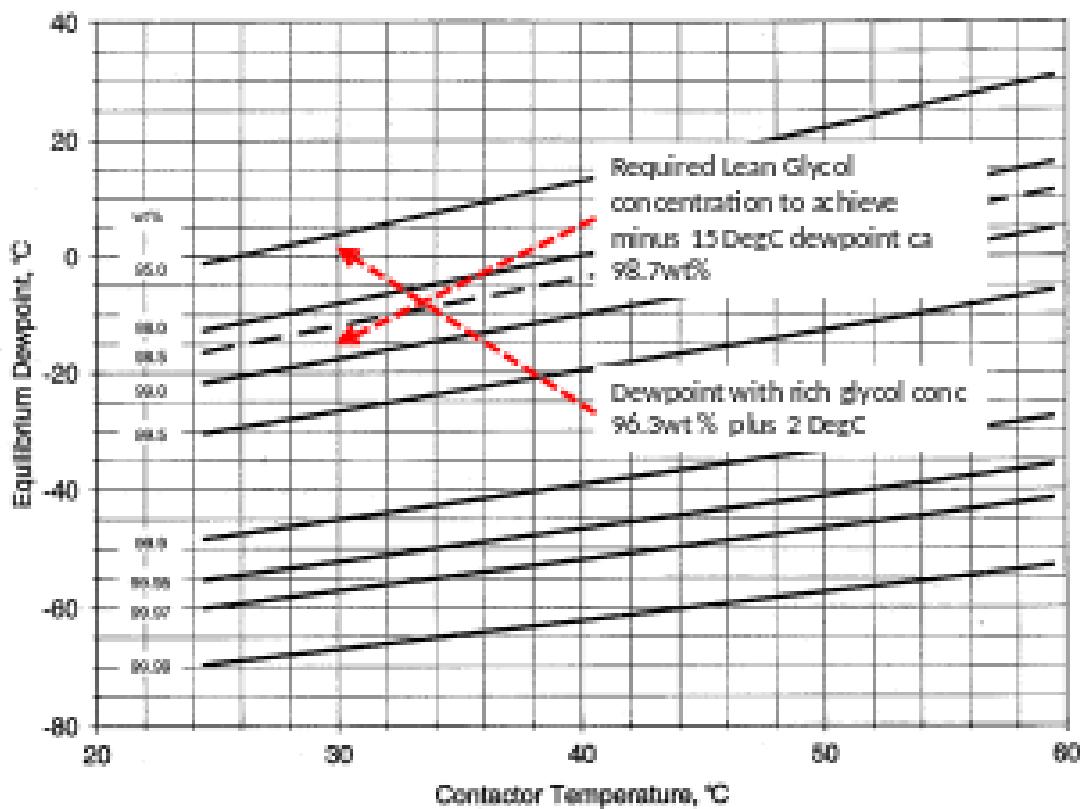
6. Use lecture notes to estimate a glycol rate (the notes on the “Glycol Circulation Rate” indicate that beyond 0.03 m<sup>3</sup> (kg H<sub>2</sub>O)<sup>-1</sup> there is no significant benefit).
7. Calculate total glycol rate,

$$Q_{glycol} = 0.0405 \frac{\text{kg H}_2\text{O}}{\text{s}} \cdot 0.03 \frac{\text{m}^3 \text{ glycol}}{\text{kg H}_2\text{O}} \quad (163)$$

$$= 0.00122 \frac{\text{m}^3 \text{ glycol}}{\text{s}} = 1.35 \frac{\text{kg glycol}}{\text{s}} \quad (164)$$

8. Calculate the outlet concentration of glycol having absorbed 0.0405 kg s<sup>-1</sup> 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<sup>-3</sup>. 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).





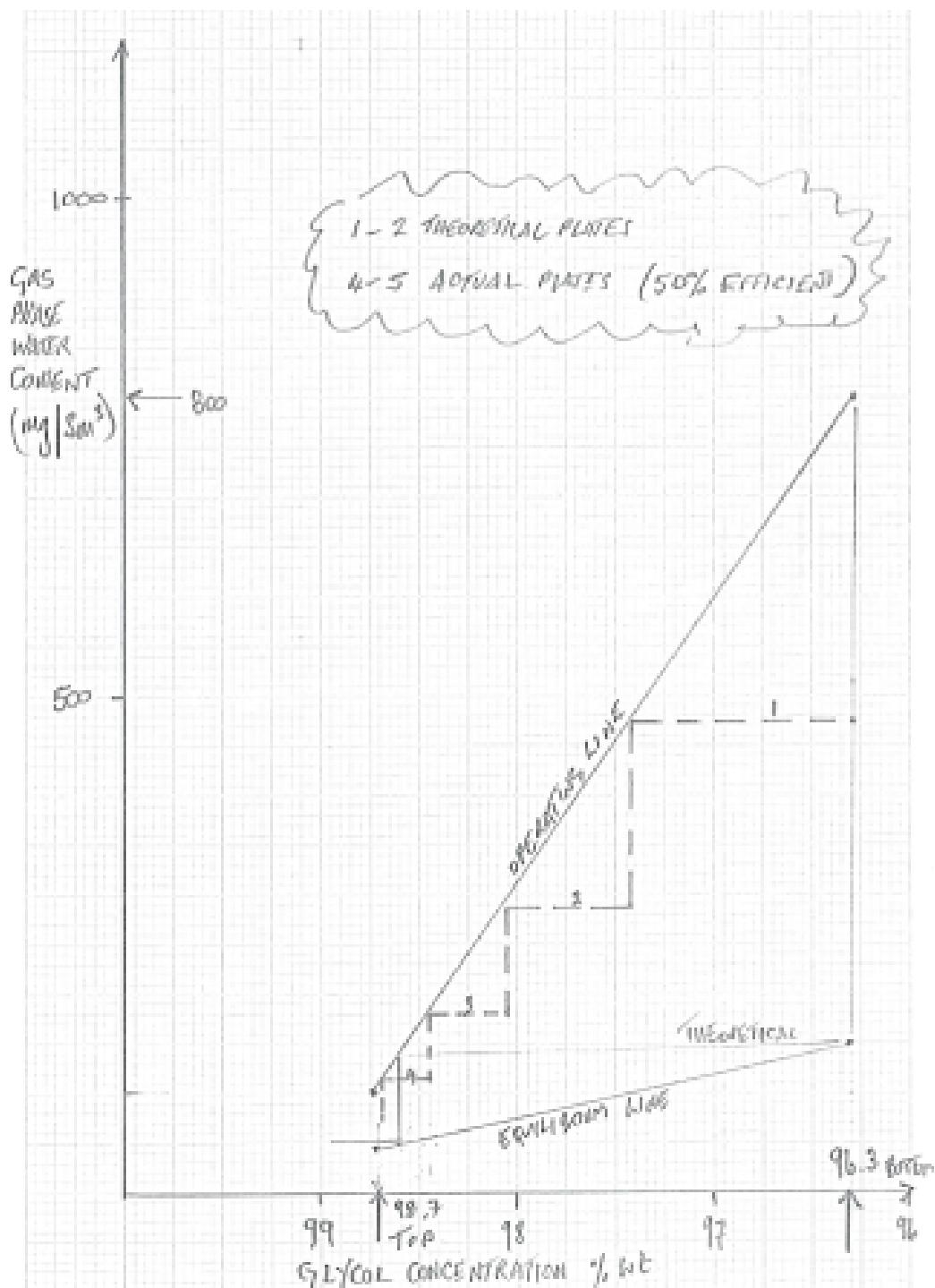
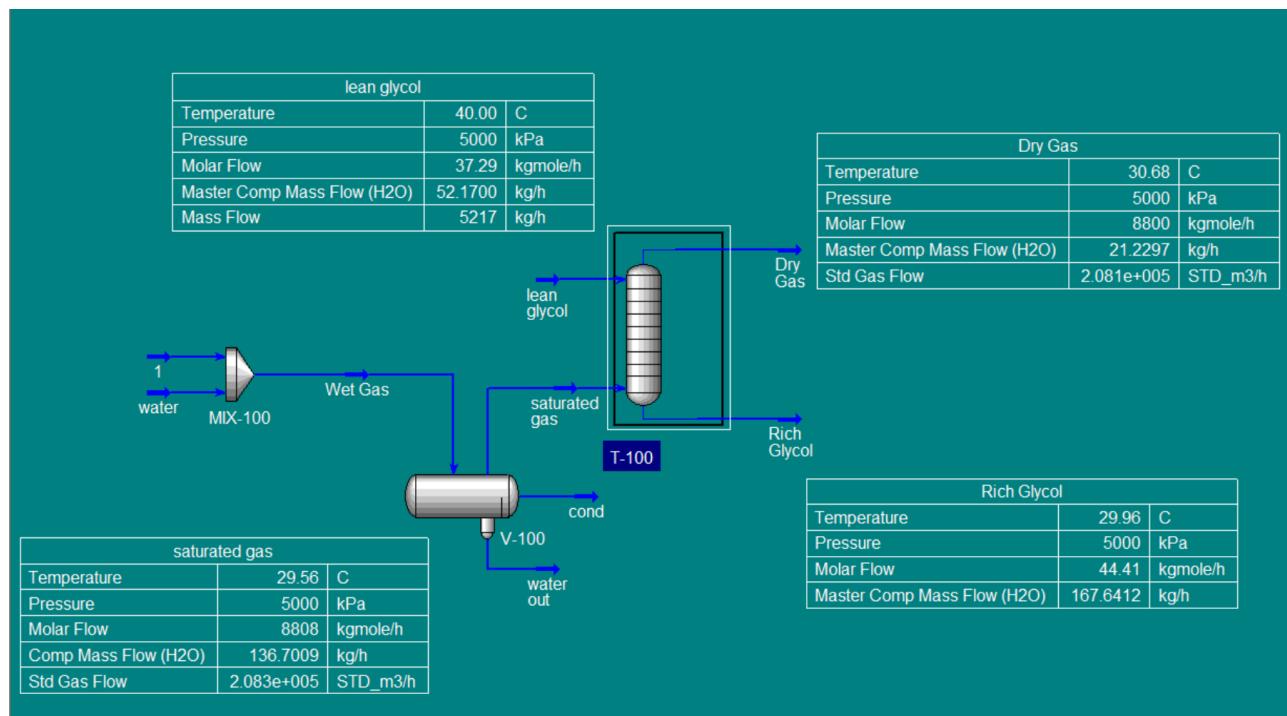


Figure 3: This graph uses an incorrect bottoms concentration of 96.3%, when it should be 95.8% but the values do not change significantly.

[Question end]

**Q.28****Question 28**

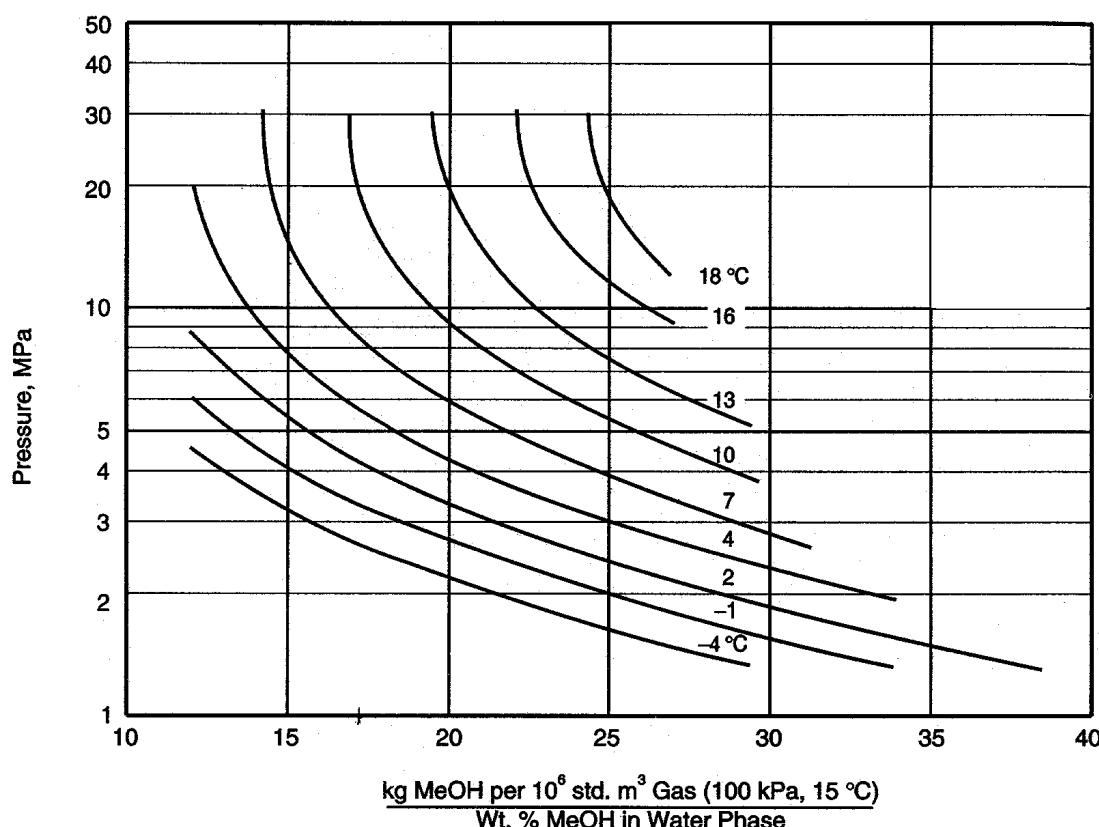
Build a glycol model on UNISIM, ASPEN HYSYS, etc. with the following conditions and compare the answers.



**[Question end]**

**Q.29****Question 29**

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  $17^\circ\text{C}$  cools to  $4.4^\circ\text{C}$  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  $30^\circ\text{C}$ .



### Solution:

Calculate the hydrate point depression

$$\Delta T = 17 \text{ } ^\circ\text{C} - 4.4 \text{ } ^\circ\text{C} = 12.6 \text{ } ^\circ\text{C} \quad (165)$$

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  $\text{H}_2\text{O}$  is therefore

$$Q_{\text{liqH}_2\text{O}} = 400 \frac{\text{kg H}_2\text{O}}{10^6 \text{ sm}^3} \cdot 0.283 \times 10^6 \frac{\text{sm}^3}{\text{day}} = 113 \frac{\text{kg H}_2\text{O}}{\text{day}} \quad (166)$$

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

$$x_{\text{MeOH}} = 100 \cdot \frac{12.6 \cdot 32}{1297 + 12.6 \cdot 32} = 23.7 \text{ wt\%} \quad (167)$$

Calculate the required mass of methanol in water

$$m_{\text{MeOH}} = 113 \frac{\text{kg H}_2\text{O}}{\text{day}} \cdot \frac{23.7 \text{ wt\% H}_2\text{O}}{100 - 23.7 \text{ wt\% H}_2\text{O}} = 35.1 \frac{\text{kg MeOH}}{\text{day}} \quad (168)$$

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

$$\text{MeOH Vap. losses (from chart)} = 15 \frac{\text{kg MeOH}/10^6 \text{sm}^3}{\text{wt\% MeOH in water}} \quad (169)$$

Use this information to calculate the mass rate of methanol lost

$$m_{\text{MeOH}} = 15 \frac{\text{kg MeOH}/10^6 \text{sm}^3}{\text{wt\% MeOH in water}} \cdot 0.283 \times 10^6 \frac{\text{sm}^3}{\text{day}} \cdot 23.7 \text{ wt\% MeOH} \quad (170)$$

$$= 100.6 \frac{\text{kg MeOH}}{\text{day}} \quad (171)$$

The total required injection rate of methanol is therefore

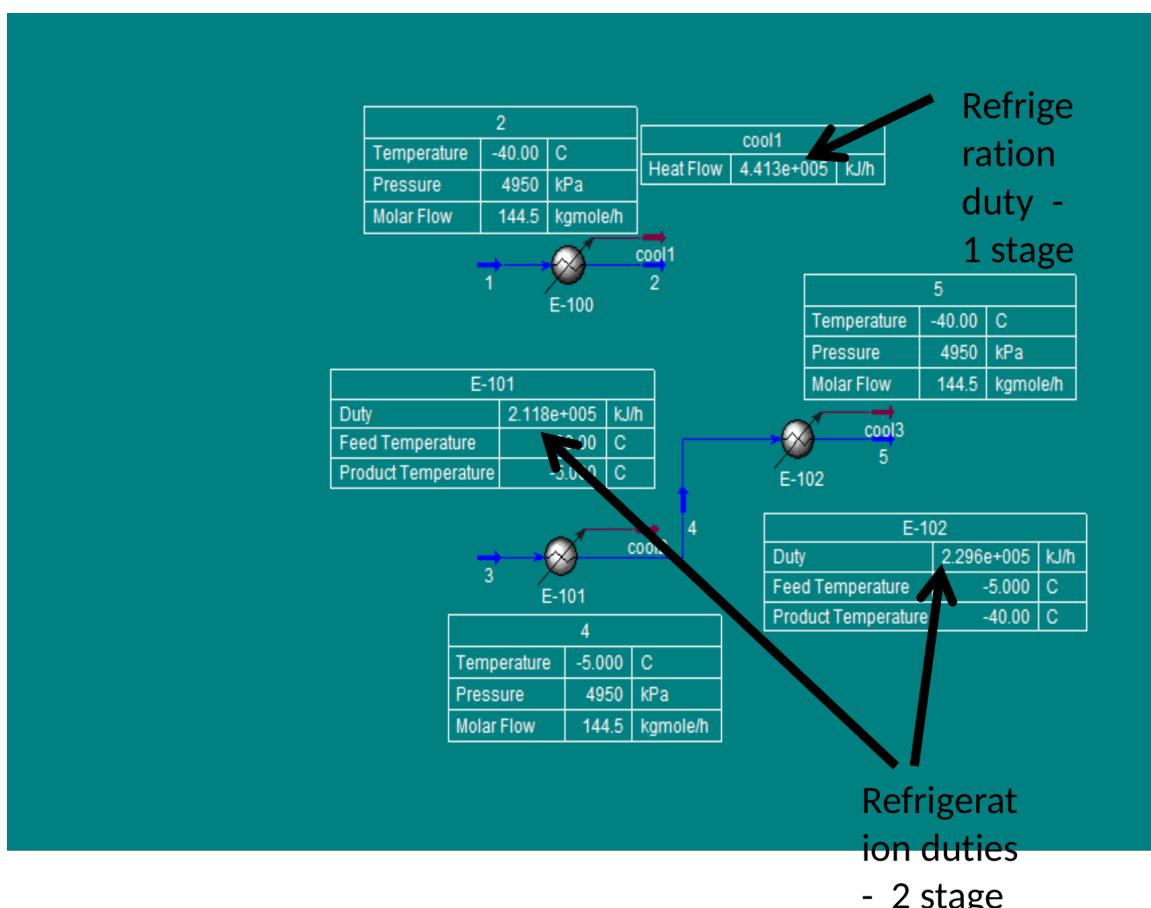
$$m_{\text{MeOH},\text{total}} = 35.1 \frac{\text{kg MeOH}}{\text{day}} + 100.6 \frac{\text{kg MeOH}}{\text{day}} = 135.7 \frac{\text{kg MeOH}}{\text{day}} \quad (172)$$

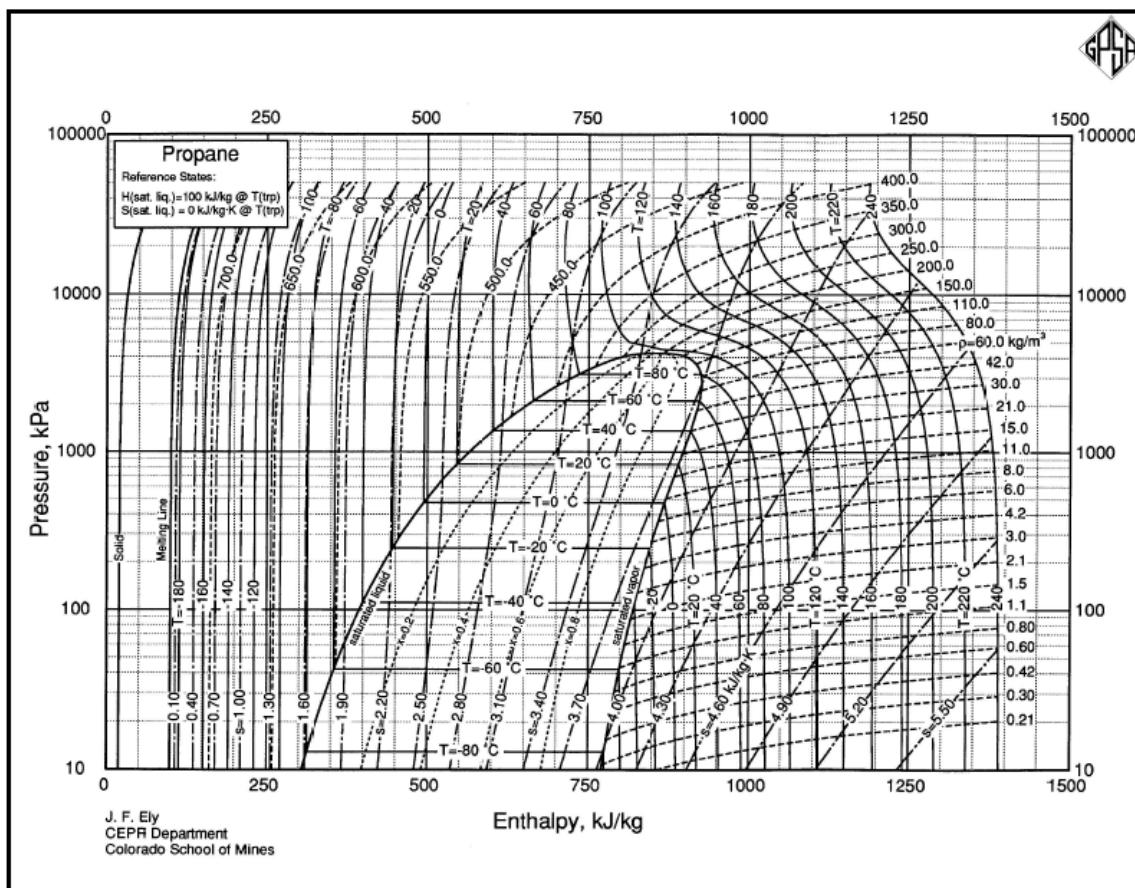
**[Question end]**

### Q.30 Question 30

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 %.





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

$$825 \frac{\text{kJ}}{\text{kg}} + 150 \frac{\text{kJ}}{\text{kg}} = 975 \frac{\text{kJ}}{\text{kg}} \quad (175)$$

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

$$570 \frac{\text{kJ}}{\text{kg}} + 825 \frac{\text{kJ}}{\text{kg}} = -255 \frac{\text{kJ}}{\text{kg}} \quad (176)$$

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

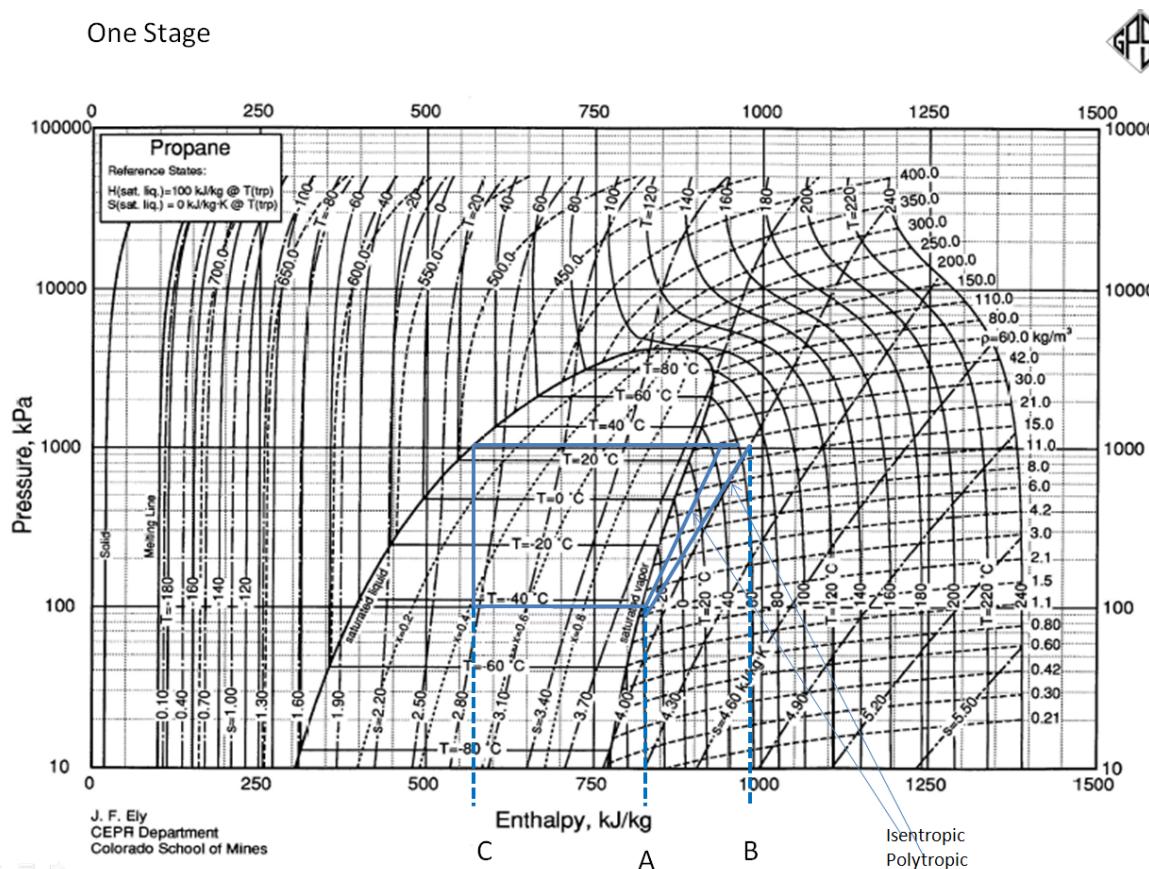
$$m_{\text{propane}} = \frac{441300 \frac{\text{kJ}}{\text{hr}}}{255 \frac{\text{kJ}}{\text{kg}}} = 1730 \frac{\text{kg}}{\text{hr}} \quad (177)$$

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

$$W_{\text{comp}} = m_{\text{propane}} \Delta H_{\text{real}} = 1730 \frac{\text{kg}}{\text{hr}} \cdot 150 \frac{\text{kJ}}{\text{kg}} = 260000 \frac{\text{kJ}}{\text{hr}} \quad (178)$$

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

$$W_{\text{cond}} = \left( 975 \frac{\text{kJ}}{\text{kg}} - 570 \frac{\text{kJ}}{\text{kg}} \right) \cdot 1730 \frac{\text{kg}}{\text{hr}} = 700000 \frac{\text{kJ}}{\text{hr}} \quad (179)$$



### Q.31 Question 31

**Past exam question**

- a) A process gas is to be cooled from  $20^\circ\text{C}$  to  $-35^\circ\text{C}$  using a refrigeration plant. The refrigerant is propane. Using the attached Mollier chart (see Fig. 4) for propane estimate the compressor energy savings between a one and a two stage refrigeration system. Assume air is available at  $25^\circ\text{C}$  for condensation. State any other assumptions. Enclose your Mollier chart including your student number with the returned exam booklet. **[14 marks]**

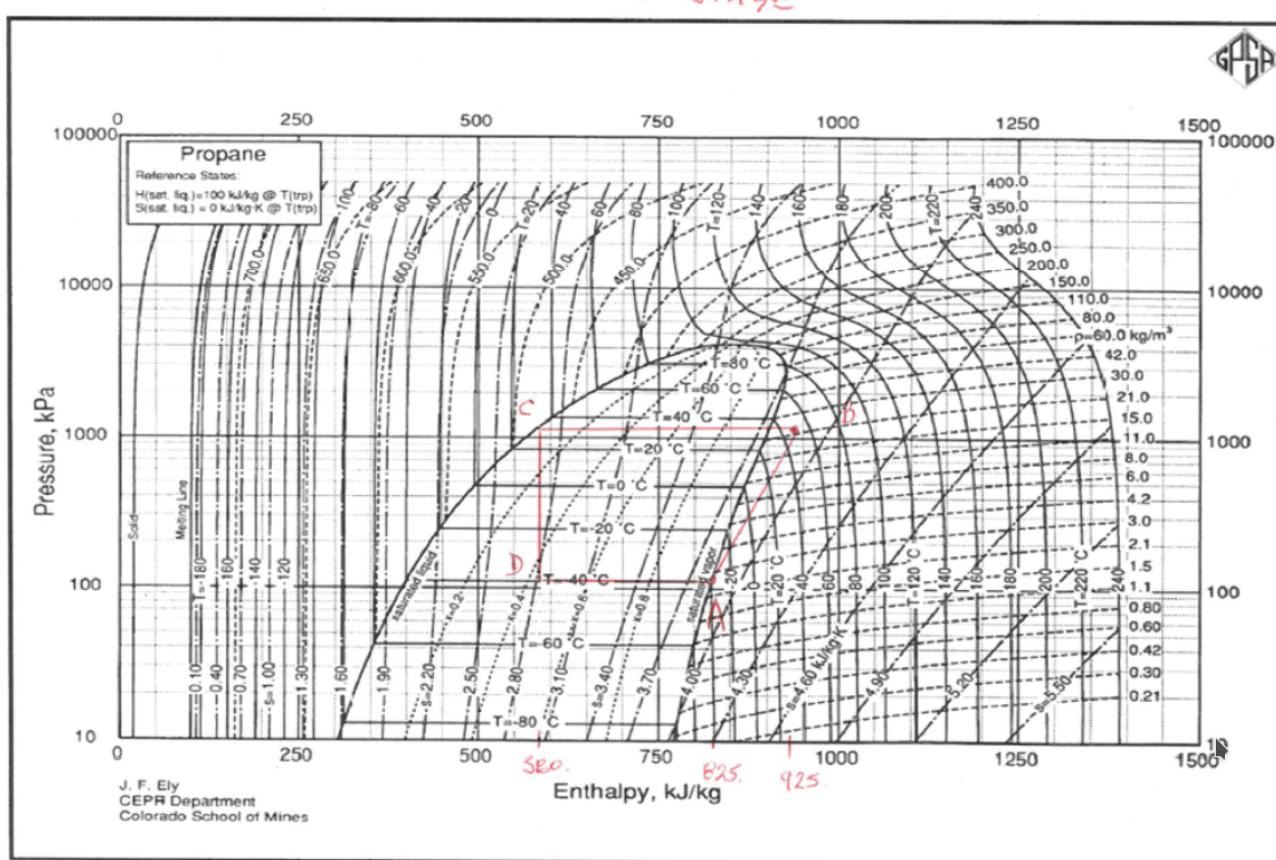
**Solution:**

Assume a  $5^\circ\text{C}$  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  $-35^\circ\text{C}$  requires an evaporator pressure of 120 kPa ( $-40^\circ\text{C}$ ). The Mollier chart can now be used to complete the one stage refrigeration system<sup>4</sup>.

**[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  $1 \text{ kJ s}^{-1}$  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  $-7.5^{\circ}\text{C}$ .

To cool the process to  $-7.5^{\circ}\text{C}$  requires evaporator pressure of 350 kPa ( $-12.5^{\circ}\text{C}$ ). The Mollier chart can now be used to complete the first stage of the two stage refrigeration system<sub>2</sub>.

[2/14] A to B compression – isentropic enthalpy change =  $925 - 850 = 75 \text{ kJ kg}^{-1}$  ✓.

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

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

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

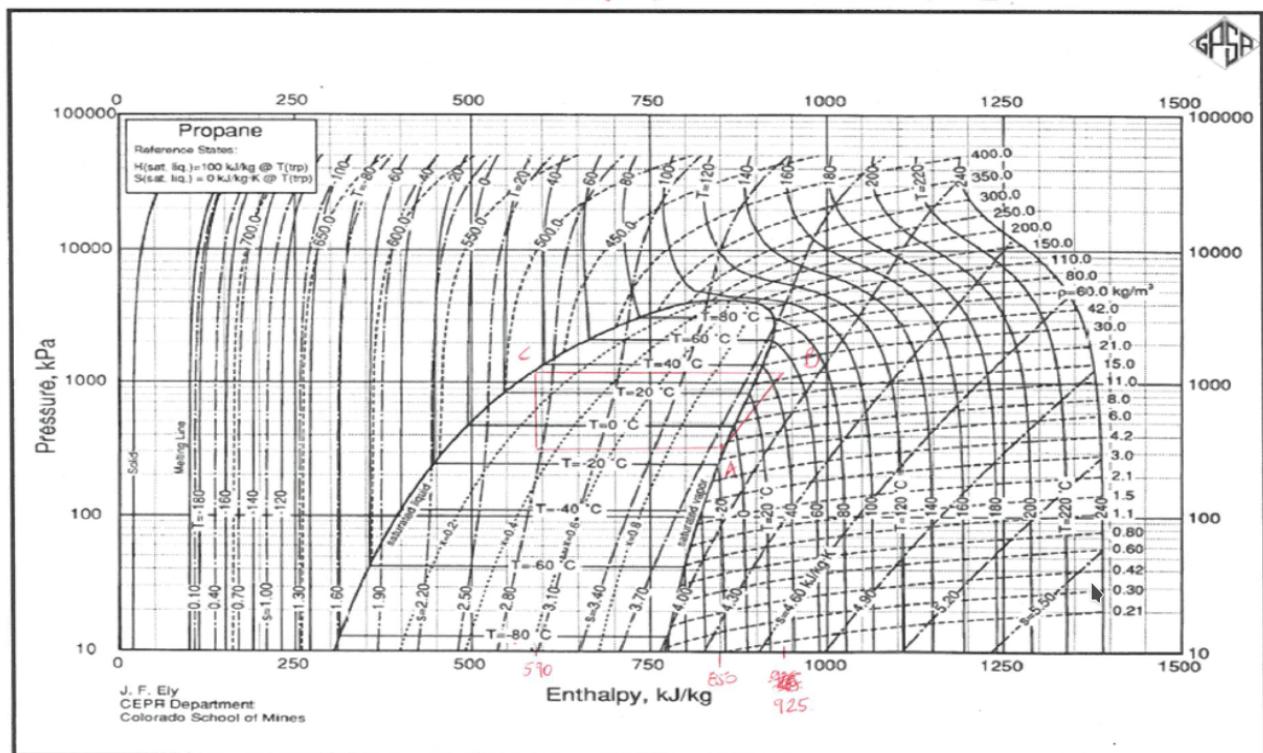
Hence compressor power =  $0.0019 \times 75 = 0.085 \text{ kJ s}^{-1}$ .

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

[1/14] Total power for two stages =  $0.205 + 0.085 = 0.29 \text{ kJ s}^{-1}$  ✓.

[1/14] So 2 stages is  $0.29/0.41 \times 100 = 70\%$  of 1 stage power<sub>1</sub> ✓.

2 STAGE - 1STAGE CYCLE



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

**Solution:**

[6/6]

(1 mark each)<sub>6</sub>

- (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.  
 Student ID:

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### 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<sup>-1</sup> 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<sup>-1</sup> (point E)
5. The enthalpy change over the compressor is therefore

$$\Delta H_{real} = 940 \frac{\text{kJ}}{\text{kg}} - 855 \frac{\text{kJ}}{\text{kg}} = 85 \frac{\text{kJ}}{\text{kg}} \quad (180)$$

6. The propane condenses at constant pressure to the same point as before at 570 kJ kg<sup>-1</sup> (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

$$570 \frac{\text{kJ}}{\text{kg}} - 855 \frac{\text{kJ}}{\text{kg}} = -285 \frac{\text{kJ}}{\text{kg}} \quad (181)$$

gives the enthalpy that is extracted from the process

9. The required refrigeration for this cooling stage is 211800 kJ hr<sup>-1</sup>. The required propane flowrate is therefore

$$m_{propane} = \frac{211800 \frac{\text{kJ}}{\text{hr}}}{285 \frac{\text{kJ}}{\text{kg}}} = 743 \frac{\text{kg}}{\text{hr}} \quad (182)$$

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

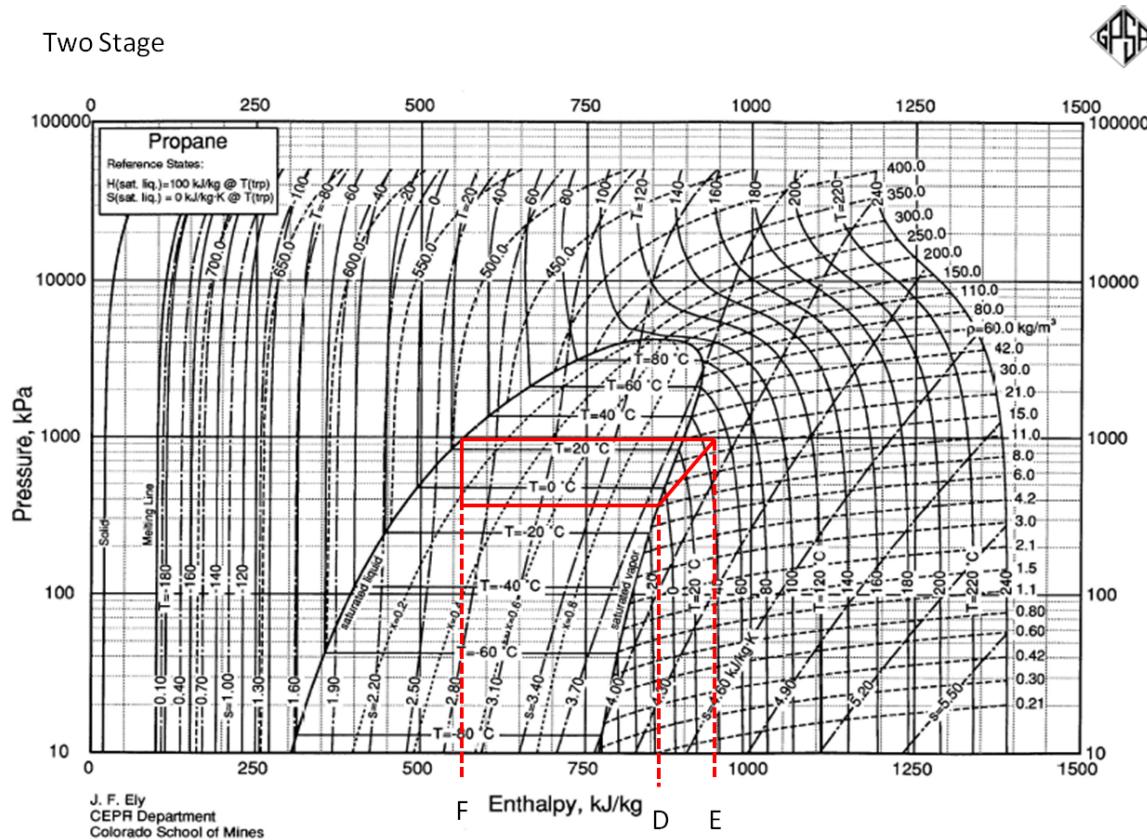
$$W_{comp} = m_{propane} \Delta H_{real} = 743 \frac{\text{kg}}{\text{hr}} \cdot 85 \frac{\text{kJ}}{\text{kg}} = 63100 \frac{\text{kJ}}{\text{hr}} \quad (183)$$

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

$$W_{cond} = \left( 940 \frac{\text{kJ}}{\text{kg}} - 570 \frac{\text{kJ}}{\text{kg}} \right) \cdot 743 \frac{\text{kg}}{\text{hr}} = 275000 \frac{\text{kJ}}{\text{hr}} \quad (184)$$

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 kJ hr<sup>-1</sup>.

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

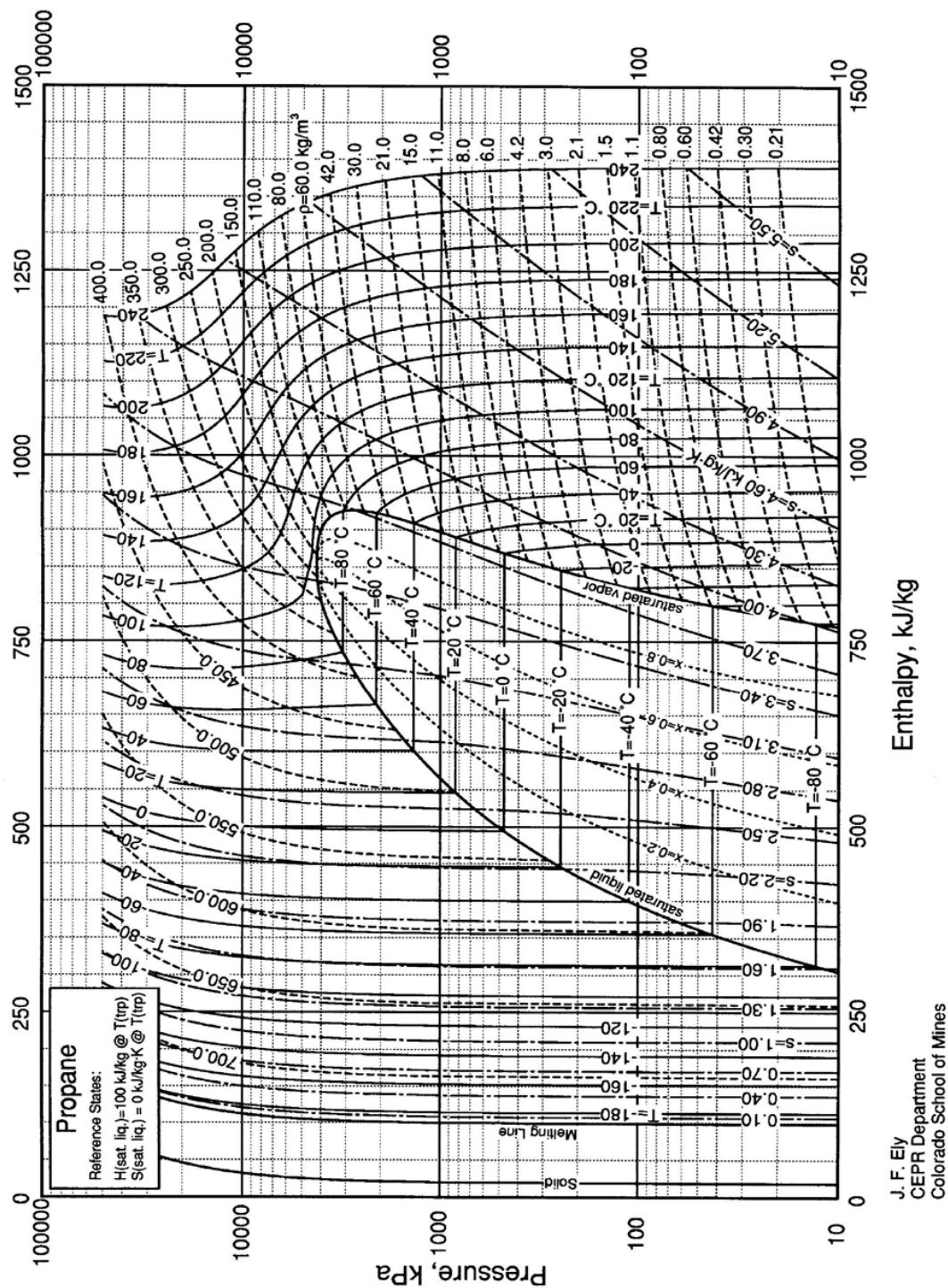


If we compare the compressor power for the 1 and 2-stage refrigeration options we find that the 2-stage compressor duty

$$W_{comp,2\text{-stage}} = 63100 \frac{\text{kJ}}{\text{hr}} + 135100 \frac{\text{kJ}}{\text{hr}} = 198100 \frac{\text{kJ}}{\text{hr}} \quad (185)$$

is 24% lower than the 1-stage duty

$$W_{comp,1\text{-stage}} = 260000 \frac{\text{kJ}}{\text{hr}} \quad (186)$$



Enthalpy, kJ/kg

J. F. Ely  
CEPR Department  
Colorado School of Mines

Figure 4: A Mollier chart for propane.

**[Question total: 20 marks]**

**Q.32 Question 32**  
**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, H<sub>2</sub>S and CO<sub>2</sub> - 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]**

- b) Hydrate formation can be calculated using K charts, an example of which is given in Fig. 5.

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.

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

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]**

- c) To prevent hydrate formation a 2366 Sm<sup>3</sup> h<sup>-1</sup> 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 m<sup>3</sup> per kg water removed is required. A safety factor of 10°C is also required.

Equilibrium data are available in Fig. 6 and 7.

**[8 marks]**

**Solution:**

From Fig. 6, saturated gas at 80 bar (8000 kPa) and 30°C has an equilibrium concentration of 550 mg Sm<sup>-3</sup>. ✓

**[3/8]**

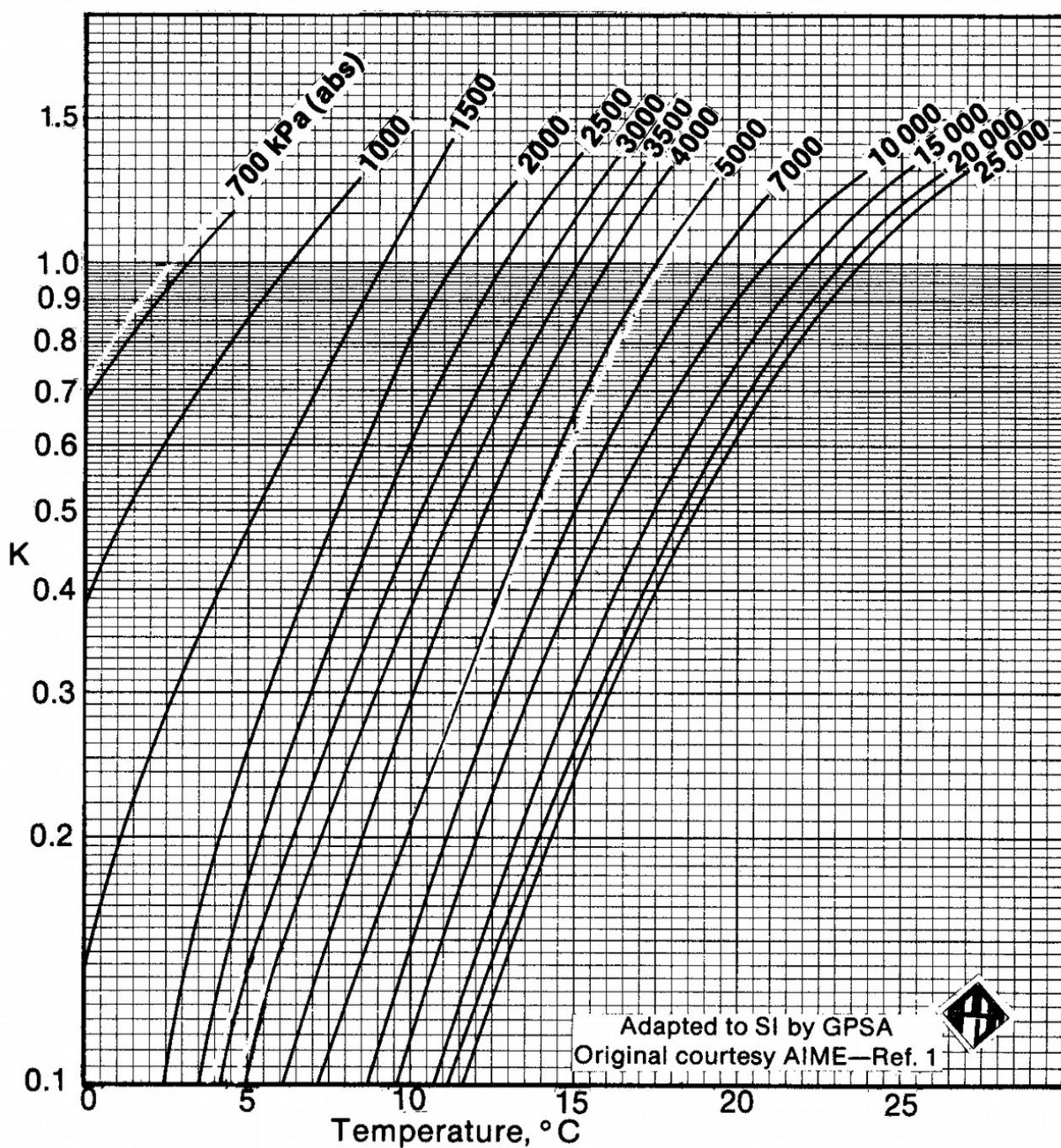


Figure 5: Hydrate K-value chart for ethane.

The glycol is available at 99% purity which, according to Fig. 7, has an equilibrium dewpoint of  $-18^{\circ}\text{C}_2$ . Unfortunately a safety factor of  $10^{\circ}\text{C}$  is needed, which implies a equilibrium of  $-8^{\circ}\text{C}$  is used for the calculations.<sup>2</sup>

Assuming little/no pressure drop in the contactor, the equilibrium wet gas concentration at 80 bar and  $-8^{\circ}\text{C}$  is  $55 \text{ mg Sm}^{-3}$  according to Fig. 6.<sup>1</sup>

The total water removal rate is  $2366(550 - 55) = 1171170 \text{ mg h}^{-1}$ , or  $1.17 \text{ kg h}^{-1}$  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  $20 \text{ mg Sm}^{-3}$ ? [2 marks]  
**Solution:**

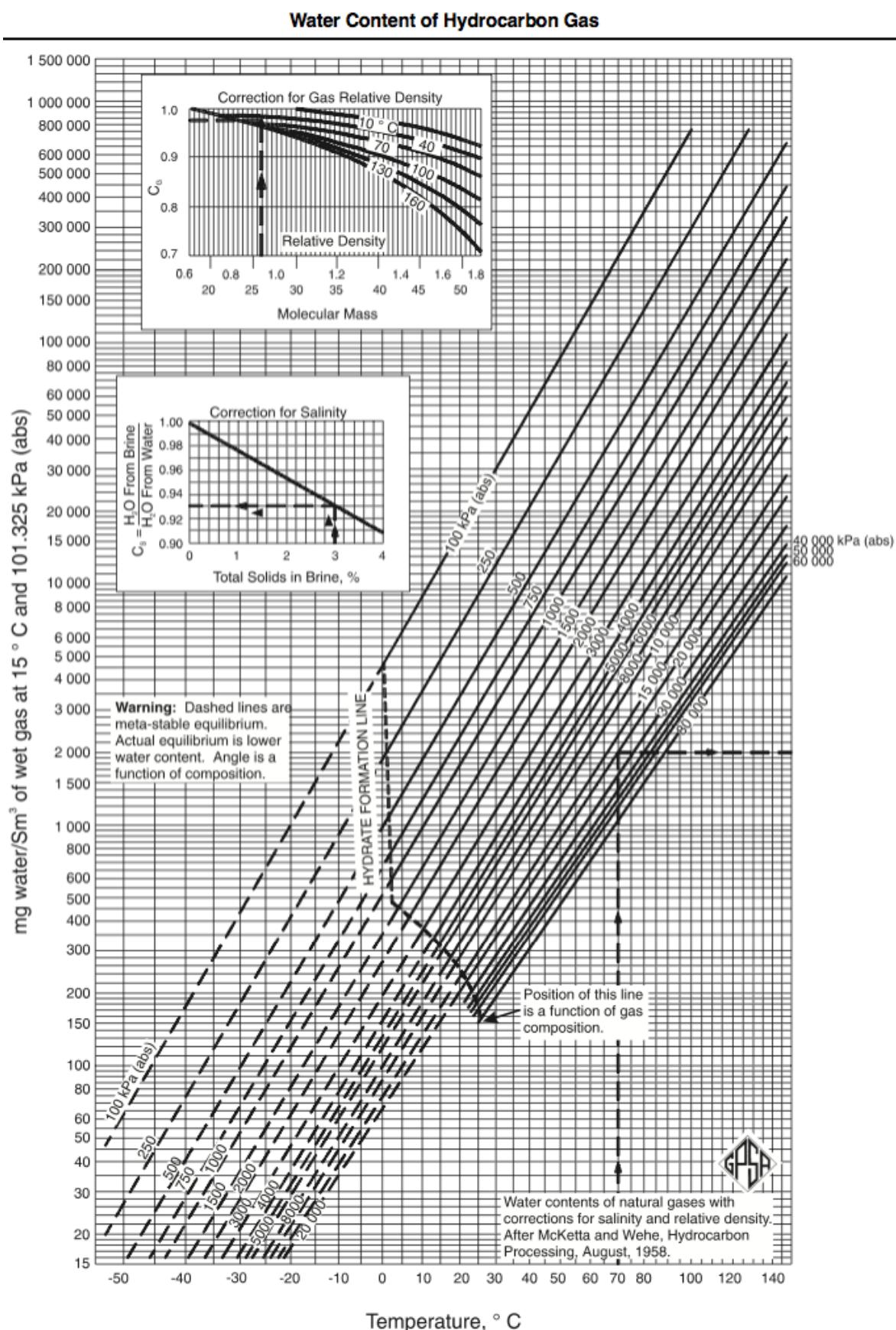


Figure 6: Equilibrium chart for wet natural gas including corrections for water salinity and relative molecular weight/density.

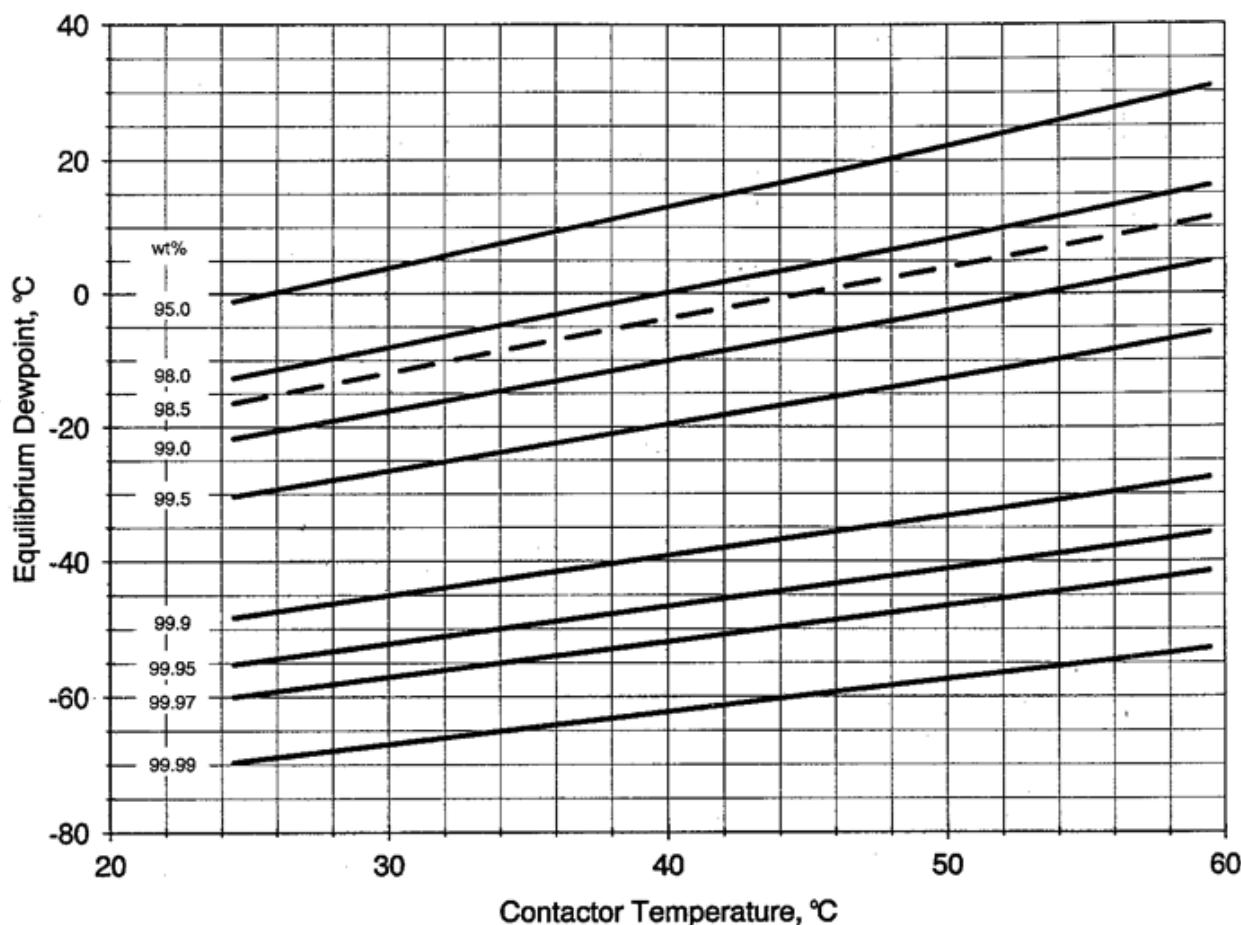


Figure 7: Equilibrium correction chart for wet natural gas over TEG solutions.

[1/2] The required dew point is  $-22.5^{\circ}\text{C}$  at 8000 kPa. With a safety margin a dewpoint of  $-32.5^{\circ}\text{C}$  is required over glycol<sup>v</sup>. At a contactor temperature of  $30^{\circ}\text{C}$ , this translates to a >99.5% but <99.9%<sup>v</sup>.

[1/2]

**[Question total: 20 marks]**

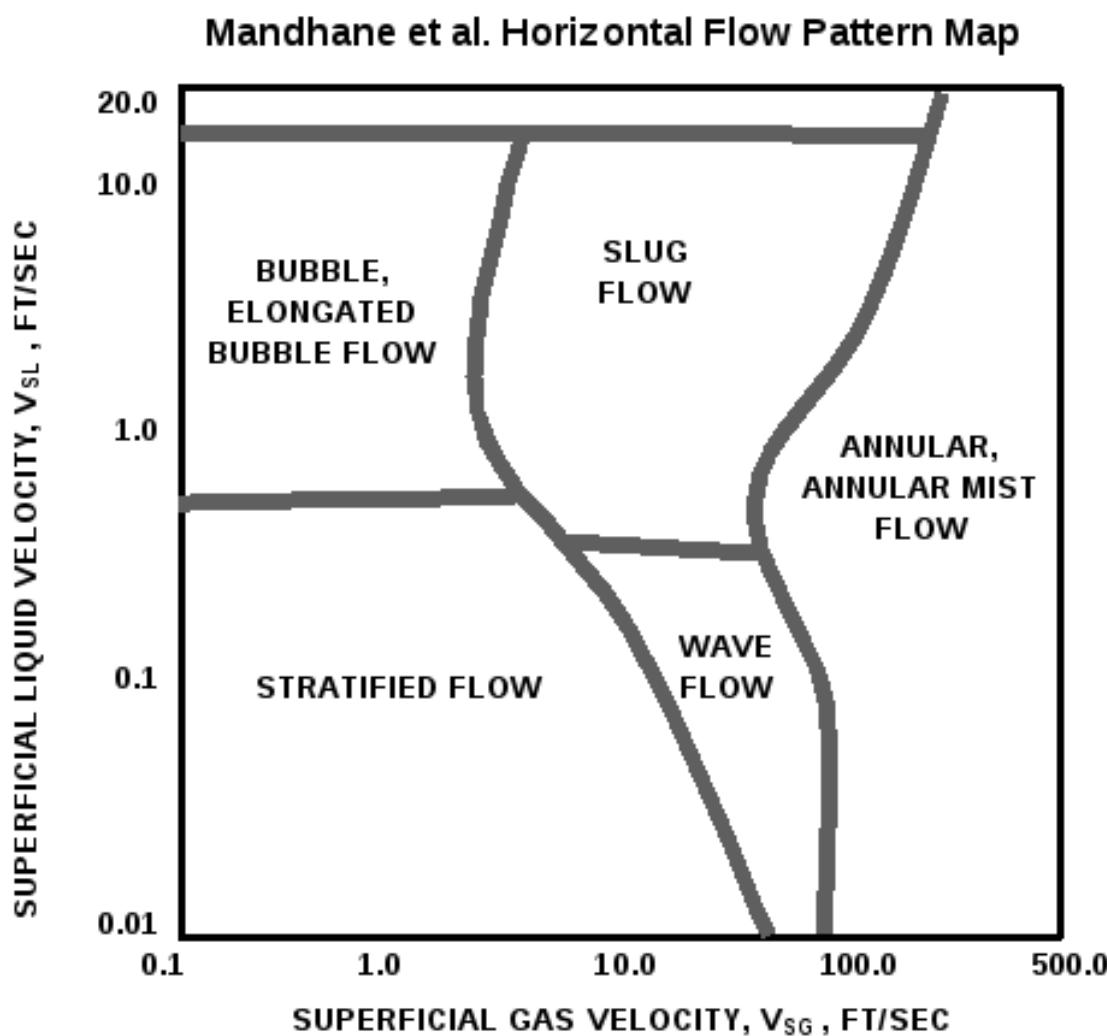
## 7 Multi-phase flow

### Q.33 Question 33

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

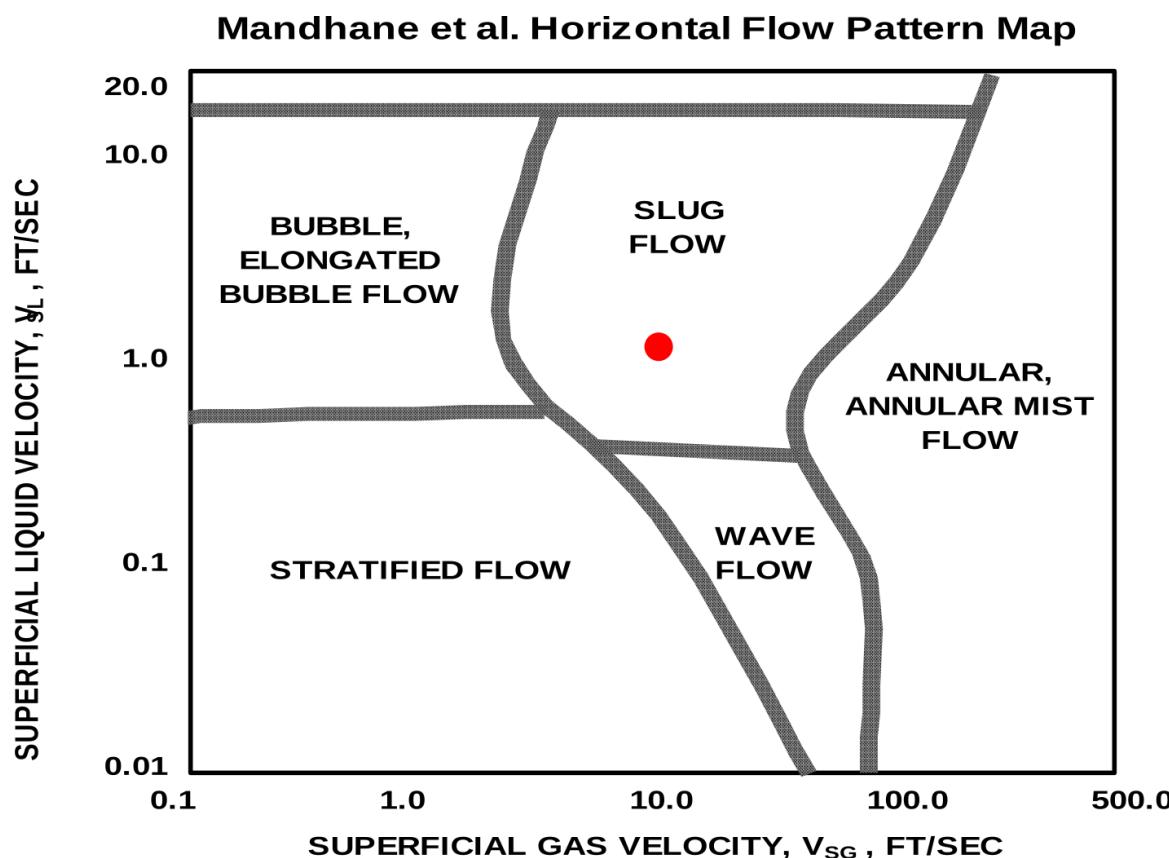
$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot 0.381 \text{ m}}{4} = 0.114 \text{ m}^2 \quad (187)$$

This is used to calculate the superficial liquid and gas velocities

$$V_{SL} = \frac{Q_L}{A} = \frac{0.0369 \text{ m}^3 \text{ s}^{-1}}{0.114 \text{ m}^2} = 0.323 \text{ m s}^{-1} = 1.06 \text{ ft s}^{-1} \quad (188)$$

$$V_{SG} = \frac{Q_G}{A} = \frac{0.835 \text{ m}^3 \text{ s}^{-1}}{0.114 \text{ m}^2} = 7.32 \text{ m s}^{-1} = 24.0 \text{ ft s}^{-1} \quad (189)$$

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

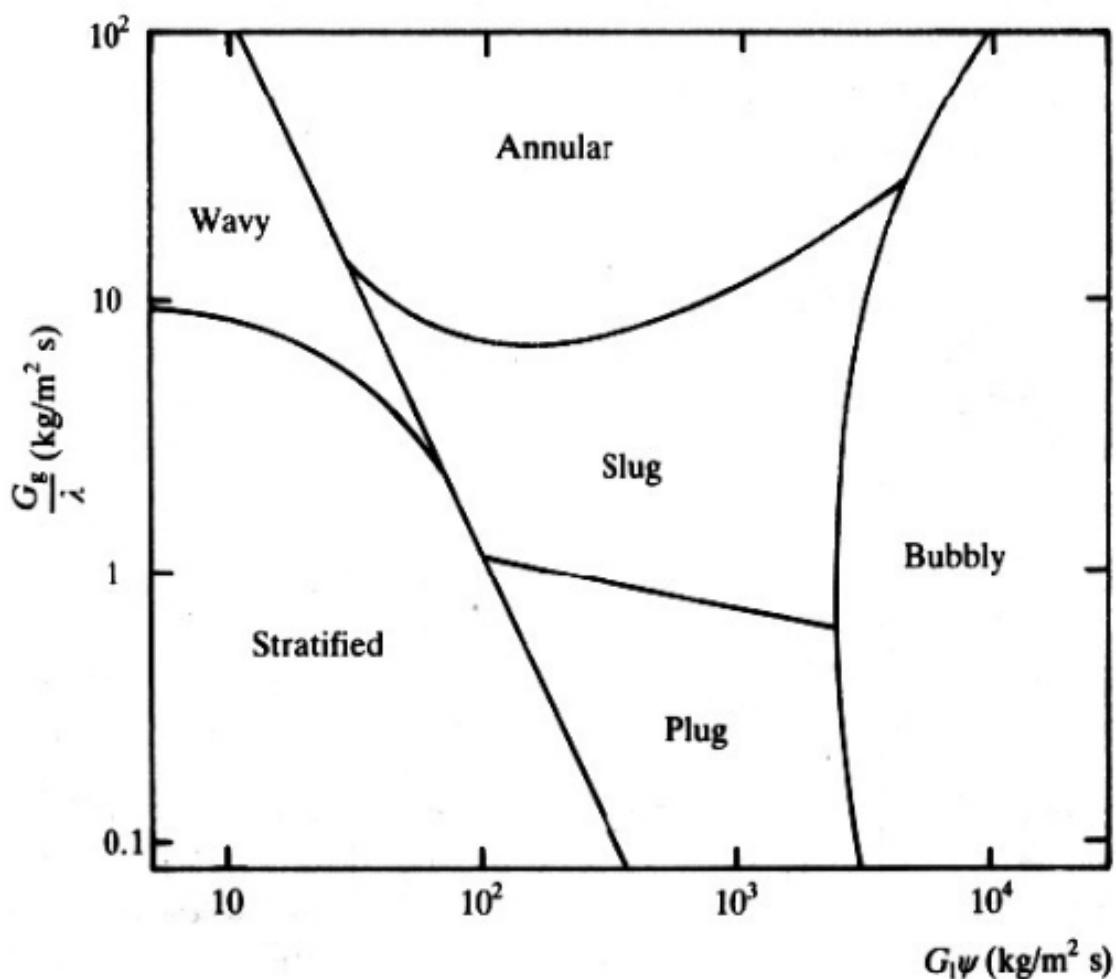


[Question end]

#### Q.34 Question 34

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 =  $0.00014 \text{ Pa s}$
- Gas viscosity =  $0.0000169 \text{ Pa s}$
- Water surface tension =  $0.029835, 0.017633, 0.0032375 \text{ N m}^{-1}$  for 30, 70, 170 bara respectively



### Solution:

Calculate the two terms  $\lambda$  and  $\psi$

$$\lambda = \sqrt{\rho'_L \rho'_G} \quad (190)$$

$$\psi = \frac{1}{\sigma'} \left( \frac{\mu'_L}{(\rho'_L)^2} \right)^{\frac{1}{3}} \quad (191)$$

where  $\sigma$  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  $Gx/\lambda$  on the y-axis and  $G(1 - x)\psi$  on the x-axis.

$p$	30	70	170	
$\sigma_l$	0.029835	0.017633	0.0032375	N/M
$\rho_l$	817	741	556	kg/m <sup>3</sup>
$\rho_g$	15	36.5	119.3	kg/m <sup>3</sup>
$\lambda$	3.19570023625496	4.74749934175877	7.43476070002705	
$\sigma'$	0.408698630136986	0.241547945205479	0.044349315068493	
$\Psi$	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	
$x$	$Gx/\lambda$			
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.35 Question 35

- 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  $Q_L = 17$  m<sup>3</sup> hr<sup>-1</sup>
  - Vapour flowrate  $Q_G = 425$  m<sup>3</sup> hr<sup>-1</sup>
  - Liquid density  $\rho_L = 880$  kg m<sup>-3</sup>
  - Gas density  $\rho_G = 20.8$  kg m<sup>-3</sup>
  - Surface tension  $\sigma = 1.5 \times 10^{-6}$  N m<sup>-1</sup>
  - 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 m<sup>3</sup>.
- The pipeline terminates with a slugcatcher to capture liquids when the pipe is sphered. The liquid capacity of the slugcatcher is 50 m<sup>3</sup>. How frequently will the line require sphering to remain within the liquid capacity?

**Solution:**

Calculate the pipe area

$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot (0.25 \text{ m})^2}{4} = 0.049 \text{ m}^2 \quad (192)$$

Calculate the superficial liquid and gas velocities

$$V_{SL} = \frac{Q_L}{A} = \frac{17 \frac{\text{m}^3}{\text{hr}}}{0.049 \text{ m}^2 \cdot 3600 \frac{\text{s}}{\text{hr}}} = 0.096 \text{ m s}^{-1} \quad (193)$$

$$V_{SG} = \frac{Q_G}{A} = \frac{425 \frac{\text{m}^3}{\text{hr}}}{0.049 \text{ m}^2 \cdot 3600 \frac{\text{s}}{\text{hr}}} = 2.41 \text{ m s}^{-1} \quad (194)$$

Calculate dimensionless groups

$$N_{LV} = 0.0565 V_{SL} \left( \frac{\rho_L}{\sigma} \right)^{0.25} = 0.846 \quad (195)$$

$$N_{GV} = 0.0565 V_{SG} \left( \frac{\rho_L}{\sigma} \right)^{0.25} = 21.1 \quad (196)$$

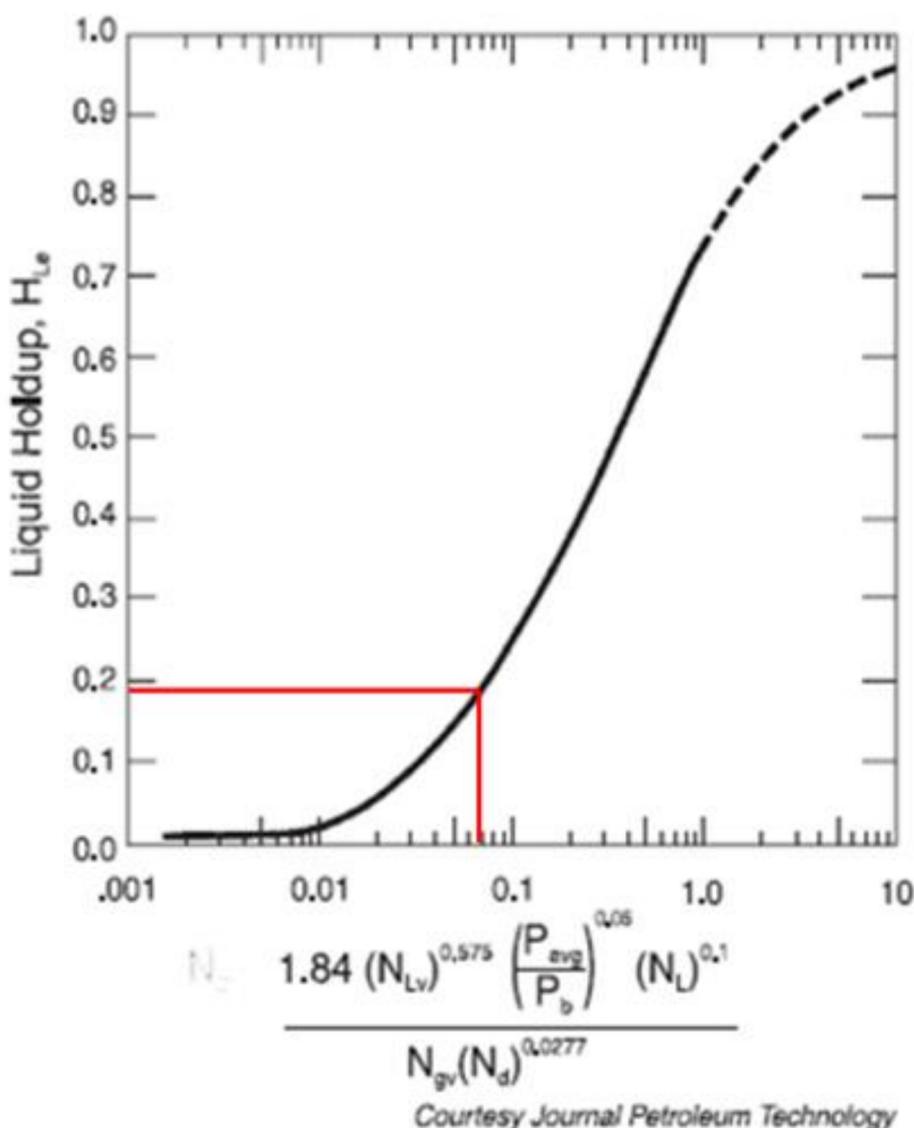
$$N_{GV} = 0.00003134 d \left( \frac{\rho_L}{\sigma} \right)^{0.5} = 190 \quad (197)$$

$$N_{GV} = 0.001769 \mu_L \left( \frac{1}{\rho_L \sigma^3} \right)^{0.25} = 0.15 \quad (198)$$

Calculate Eaton's abscissa

$$N_e = \frac{1.84 (N_{LV})^{0.575} \left( \frac{P_{avg}}{P_b} \right)^{0.05} (N_L)^{0.1}}{N_{GV} (N_d)^{0.0277}} = 0.0665 \quad (199)$$

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 m<sup>3</sup>.

The total volume of the pipe is

$$V_{pipe} = 50000 \text{ m} \cdot 0.049 \text{ m}^3 = 2450 \text{ m}^3 \quad (200)$$

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

$$\text{Total liquid hold up} = 0.19 \cdot 2450 \text{ m}^3 = 466 \text{ m}^3 \quad (201)$$

- The pipeline terminates with a slugcatcher to capture liquids when the pipe is spherized. The liquid capacity of the slugcatcher is 50 m<sup>3</sup>. 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 m<sup>3</sup>.

$$t = \frac{V_{slug\ catcher}}{Q_L} = \frac{50 \text{ m}^3}{17 \frac{\text{m}^3}{\text{hr}}} = 2.96 \text{ hr} \quad (202)$$

**[Question end]**

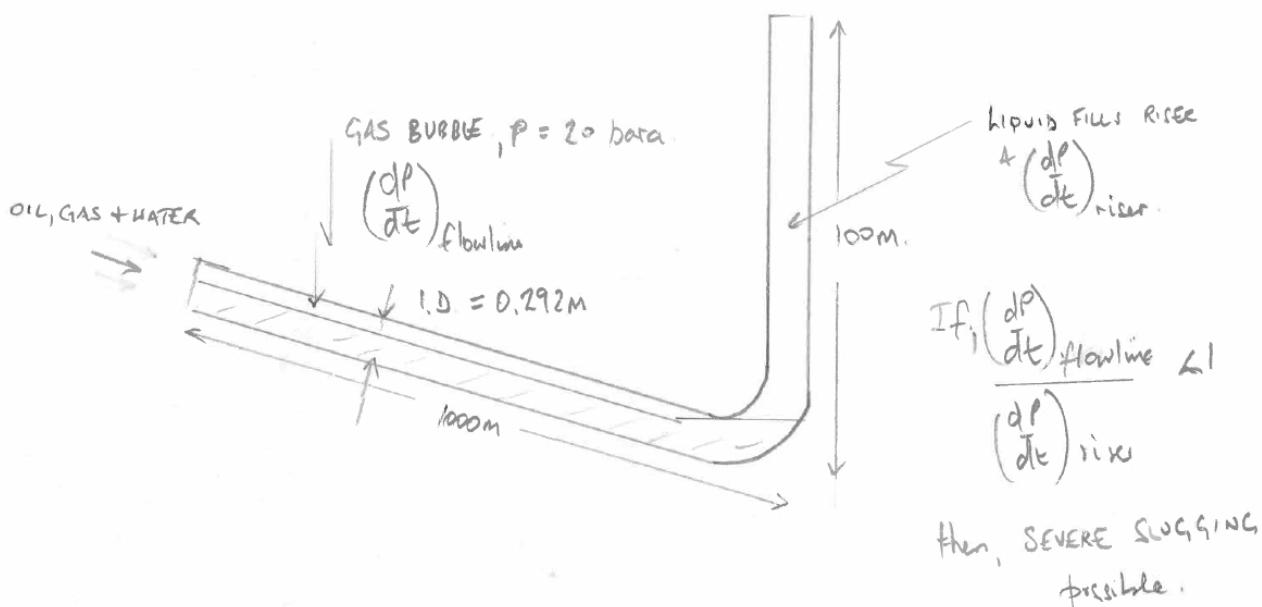
## Q.36

**Question 36**

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  $1000 \text{ kg m}^{-3}$  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

$$A = \frac{\pi D^2}{4} = \frac{\pi (0.292 \text{ m})^2}{4} = 0.067 \text{ m}^2 \quad (203)$$

The system will be prone to severe slugging if

$$\left(\frac{dp}{dt}\right)_{\text{riser}} > \left(\frac{dp}{dt}\right)_{\text{pipeline}} \quad (204)$$

**Riser**

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

$$v_{L,\text{riser}} = \frac{Q_L}{A} = \frac{0.018 \frac{\text{m}^3}{\text{s}}}{0.067 \text{ m}^2} = 0.269 \frac{\text{m}}{\text{s}} \quad (205)$$

The liquid mixture density in the riser will be

$$\rho_{mL} = \frac{Q_{oil}}{Q_{oil} + Q_{water}} \rho_{oil} + \frac{Q_{water}}{Q_{oil} + Q_{water}} \rho_{water} = 0.5 \cdot 800 \frac{\text{kg}}{\text{m}^3} + 0.5 \cdot 1000 \frac{\text{kg}}{\text{m}^3} = 900 \frac{\text{kg}}{\text{m}^3} \quad (206)$$

The rate of pressure increase in the riser will be

$$\left( \frac{dp}{dt} \right)_{\text{riser}} = \rho_L g \frac{dh}{dt} = \rho_{mL} g v_{L,\text{riser}} = 900 \frac{\text{kg}}{\text{m}^3} \cdot 9.81 \frac{\text{m}}{\text{s}^2} \cdot 0.269 \frac{\text{m}}{\text{s}} = 0.02373 \frac{\text{bar}}{\text{s}} \quad (207)$$

## Pipeline

The no-slip liquid hold up in the pipeline is

$$\lambda_L = \frac{Q_L}{Q_L + Q_G} = \frac{0.018 \frac{\text{m}^3}{\text{s}}}{0.018 \frac{\text{m}^3}{\text{s}} + 0.041 \frac{\text{m}^3}{\text{s}}} = 0.31 \quad (208)$$

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

$$V_G = A L (1 - \lambda_L) = 0.067 \text{ m}^2 \cdot 1000 \text{ m} \cdot 0.69 = 46.23 \text{ m}^3 \quad (209)$$

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

$$\left( \frac{dp}{dt} \right)_{\text{pipeline}} = \frac{dn}{dt} \frac{RT}{V_G} \quad (210)$$

The molar flowrate is

$$\frac{dn}{dt} = \frac{Q_G}{V_m} \quad (211)$$

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

$$\frac{V}{n} = V_m = \frac{RT}{P} \quad (212)$$

The rate of increase of pressure is therefore

$$\left( \frac{dp}{dt} \right)_{\text{pipeline}} = \frac{P Q_G}{V_G} = \frac{20 \text{ bar} \cdot 0.041 \frac{\text{m}^3}{\text{s}}}{46.23 \text{ m}^3} = 0.018 \frac{\text{bar}}{\text{s}} \quad (213)$$

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

$$0.02373 \frac{\text{bar}}{\text{s}} > 0.018 \frac{\text{bar}}{\text{s}} \quad (214)$$

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

**[Question end]**

Q.37

**Question 37**

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	$\text{kg m}^{-3}$
Gas density	25	$\text{kg m}^{-3}$
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	$\text{kg s}^{-1}$
Pipe i.d.	0.2	m
Pipe length	1000	m
Vapour quality	0.03	Mass fraction

**Solution:**

First, calculate the pipe cross sectional area

$$A = \frac{\pi D^2}{4} = \frac{\pi \cdot 0.2^2}{4} = 0.314 \text{ m}^2 \quad (215)$$

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

$$\Delta P = \frac{2 f_{2p} \rho_n v_m^2 L}{D} \quad (216)$$

where  $\rho_n$  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

$$Re_n = \frac{\rho_n v_m D}{\mu_n} \quad (217)$$

where  $\mu_n$  is the no-slip mixture viscosity. We, therefore, need to calculate  $v_m$ ,  $\rho_n$ , and  $\mu_n$ . Start by calculating the liquid and gas flowrates

$$Q_L = \frac{M(1-x)}{\rho_L} = \frac{50 \frac{\text{kg}}{\text{s}} (1-0.03)}{860 \frac{\text{kg}}{\text{m}^3}} = 0.056 \text{ m}^3 \text{ s}^{-1} \quad (218)$$

$$Q_G = \frac{Mx}{\rho_G} = \frac{50 \frac{\text{kg}}{\text{s}} \cdot 0.03}{25 \frac{\text{kg}}{\text{m}^3}} = 0.06 \text{ m}^3 \text{ s}^{-1} \quad (219)$$

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

$$v_{SL} = \frac{Q_L}{A} = \frac{0.056 \frac{\text{m}^3}{\text{s}}}{0.314 \text{ m}^2} = 1.796 \text{ m s}^{-1} \quad (220)$$

$$v_{SG} = \frac{Q_G}{A} = \frac{0.06 \frac{\text{m}^3}{\text{s}}}{0.314 \text{ m}^2} = 1.911 \text{ m s}^{-1} \quad (221)$$

We can now calculate the mixture velocity

$$v_m = v_{SL} + v_{SG} = 1.796 + 1.911 = 3.707 \text{ m s}^{-1} \quad (222)$$

as well as the no-slip liquid hold-up

$$\lambda_L = \frac{v_{SL}}{v_m} = \frac{1.796 \frac{\text{m}}{\text{s}}}{3.707 \frac{\text{m}}{\text{s}}} = 0.4845 \quad (223)$$

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

$$\rho_n = \lambda_L \rho_L + (1 - \lambda_L) \rho_G = 0.4845 \cdot 860 \frac{\text{kg}}{\text{m}^3} + (1 - 0.4845) \cdot 25 \frac{\text{kg}}{\text{m}^3} = 429.57 \text{ kg m}^{-3} \quad (224)$$

$$\mu_n = \lambda_L \mu_L + (1 - \lambda_L) \mu_G = 0.4845 \cdot 0.0013 \frac{\text{kg}}{\text{m s}} + (1 - 0.4845) \cdot 0.000012 \frac{\text{kg}}{\text{m s}} = 0.000636 \text{ kg m}^{-1} \text{ s}^{-1} \quad (225)$$

We now have the information available to calculate the Reynolds number

$$Re_n = \frac{\rho_n v_m D}{\mu_n} = \frac{429.57 \frac{\text{kg}}{\text{m}^3} \cdot 3.707 \frac{\text{m}}{\text{s}} \cdot 0.2 \text{ m}}{0.000636 \frac{\text{kg}}{\text{m s}}} = 500697 \quad (226)$$

We then calculate the friction factor using the Blasius equation

$$f_{2p} = 0.079 Re_n^{-0.25} = 0.079 \cdot 500697^{-0.25} = 0.00297 \quad (227)$$

The predicted pressure drop in the pipeline is therefore

$$\Delta P = \frac{2 f_{2p} \rho_n v_m^2 L}{D} = \frac{2 \cdot 0.00297 \cdot 429.57 \frac{\text{kg}}{\text{m}^3} \cdot (3.707 \frac{\text{m}}{\text{s}})^2 \cdot 1000 \text{ m}}{0.2 \text{ m}} = 1.753 \text{ bar} \quad (228)$$

**[Question end]**

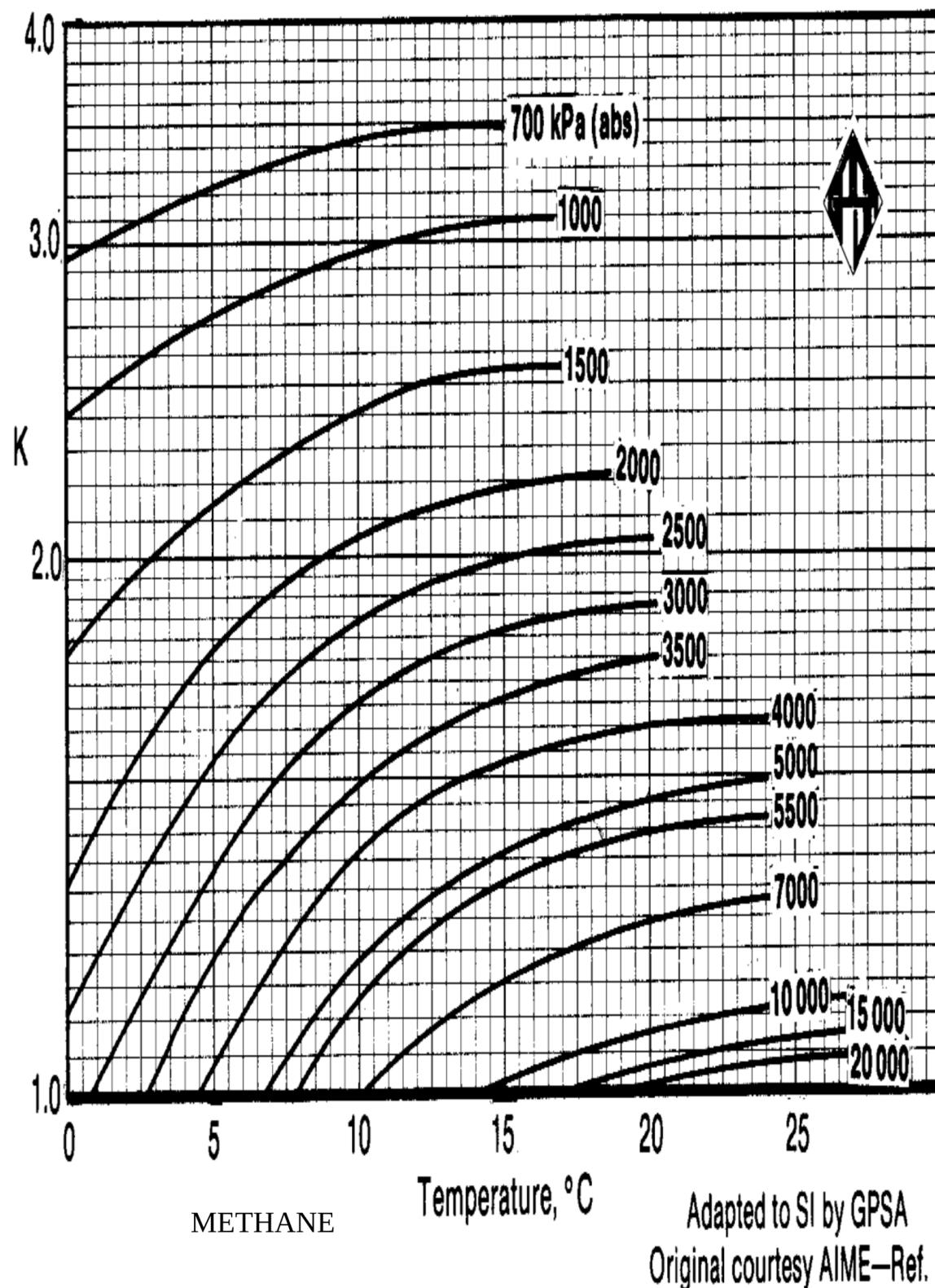
## 8 Hydrates

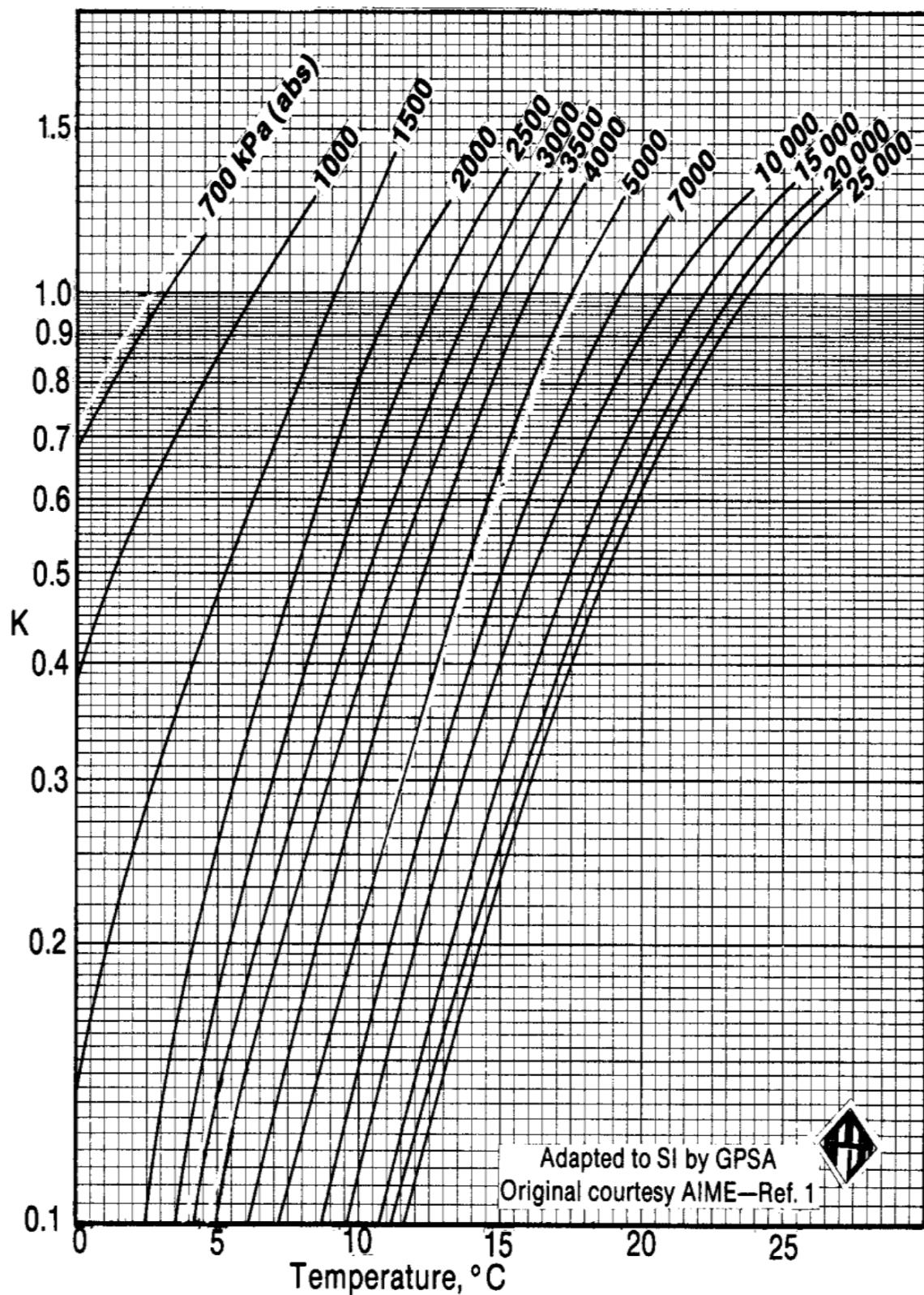
### Q.38 Question 38

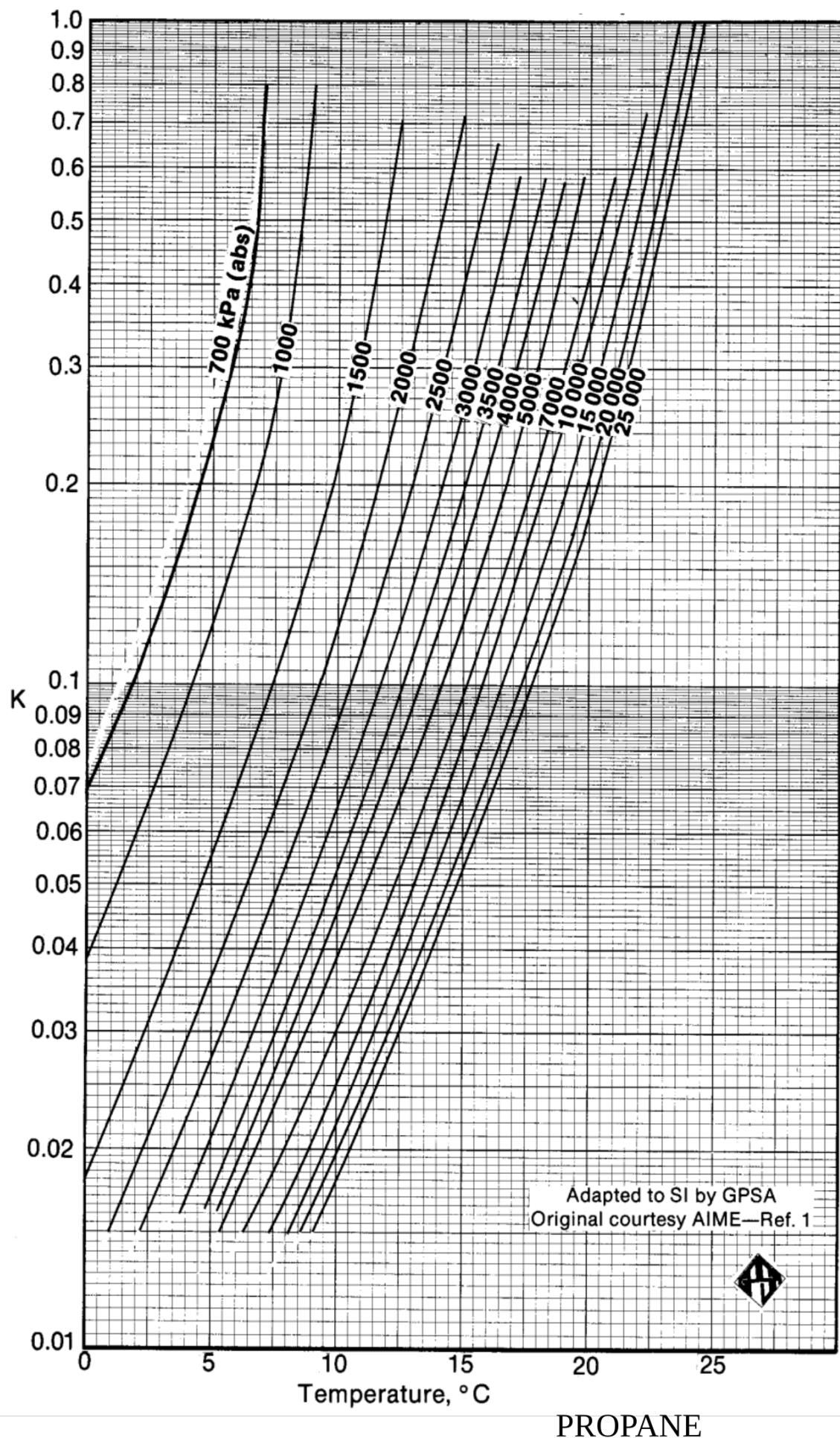
The following mixture is flowing in a subsea pipeline.

Component	Mol fraction
Methane	0.6
Ethane	0.3
Propane	0.1

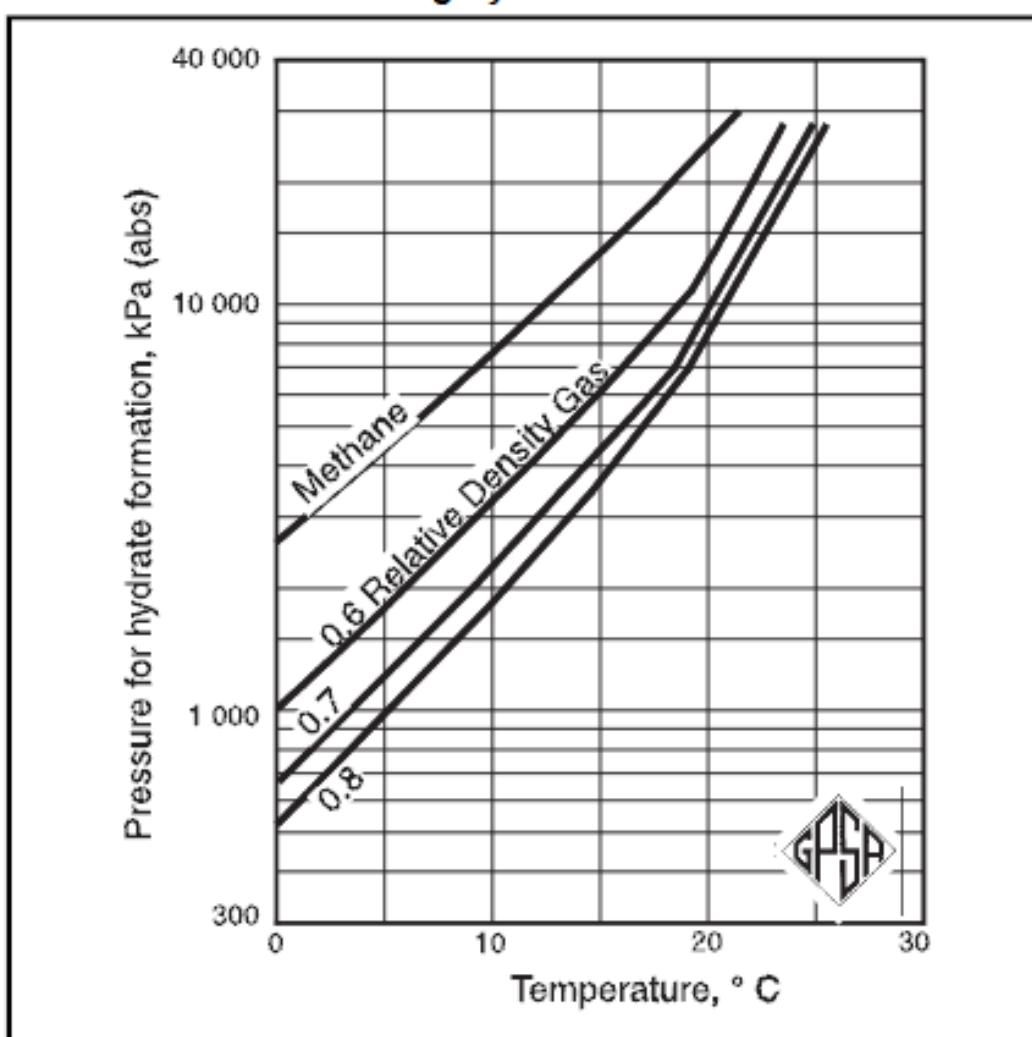
Calculate the hydrate formation temperature for a pipeline pressure of 40 bara. Use the  $K$  value figures and compare with the GPSA charts.







### Pressure-Temperature Curves for Predicting Hydrate Formation



**Solution:**

#### Using the GPSA chart

The molecular weights of the components are

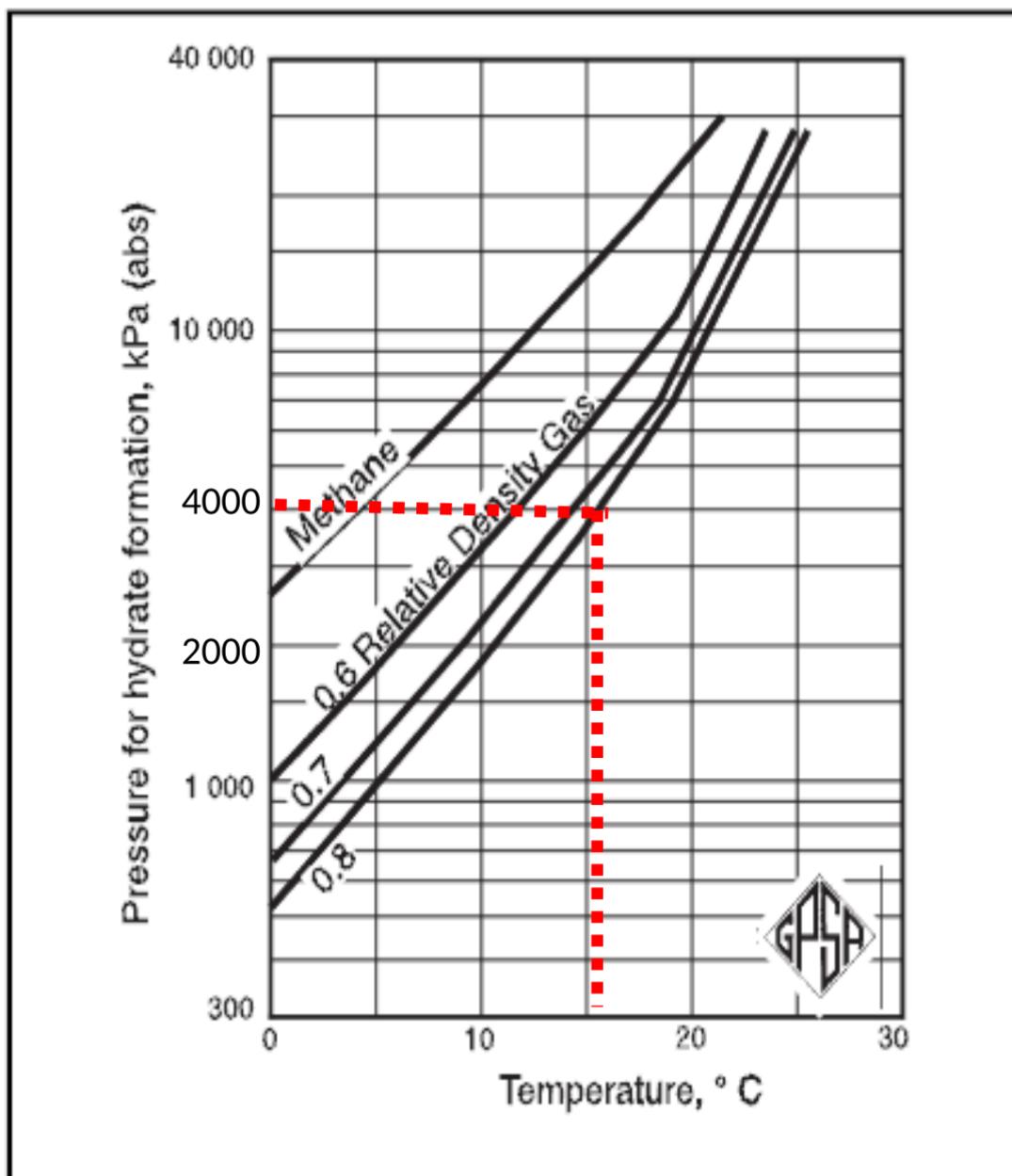
- Methane:  $16 \text{ g mol}^{-1}$
- Ethane:  $30 \text{ g mol}^{-1}$
- Propane:  $44 \text{ g mol}^{-1}$

Calculate the average molecular weight of the mixture

$$M_{\text{ave}} = 0.6 \cdot 16 \frac{\text{g}}{\text{mol}} + 0.3 \cdot 30 \frac{\text{g}}{\text{mol}} + 0.1 \cdot 44 \frac{\text{g}}{\text{mol}} = 23 \frac{\text{g}}{\text{mol}} \quad (229)$$

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



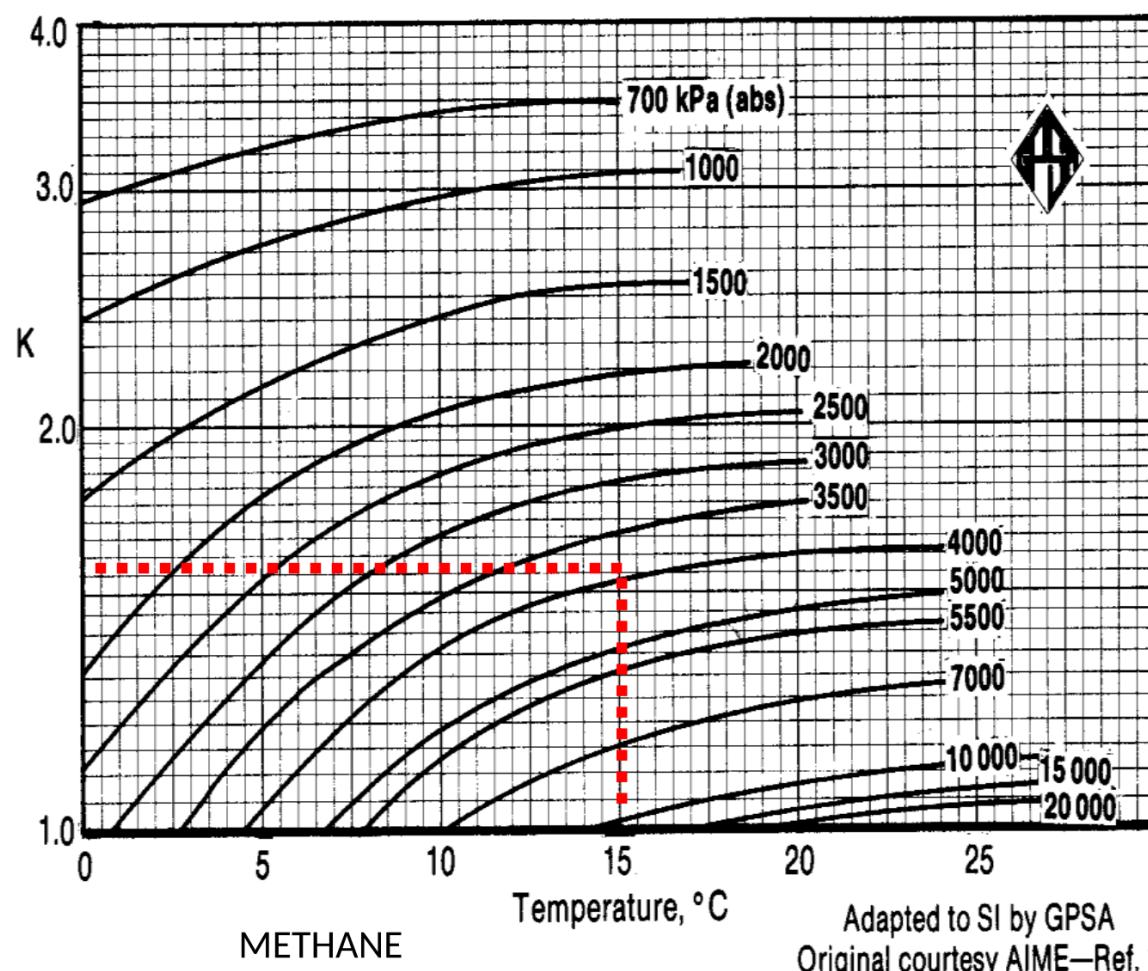
The chart gives a hydrate formation temperature of 16 °C.

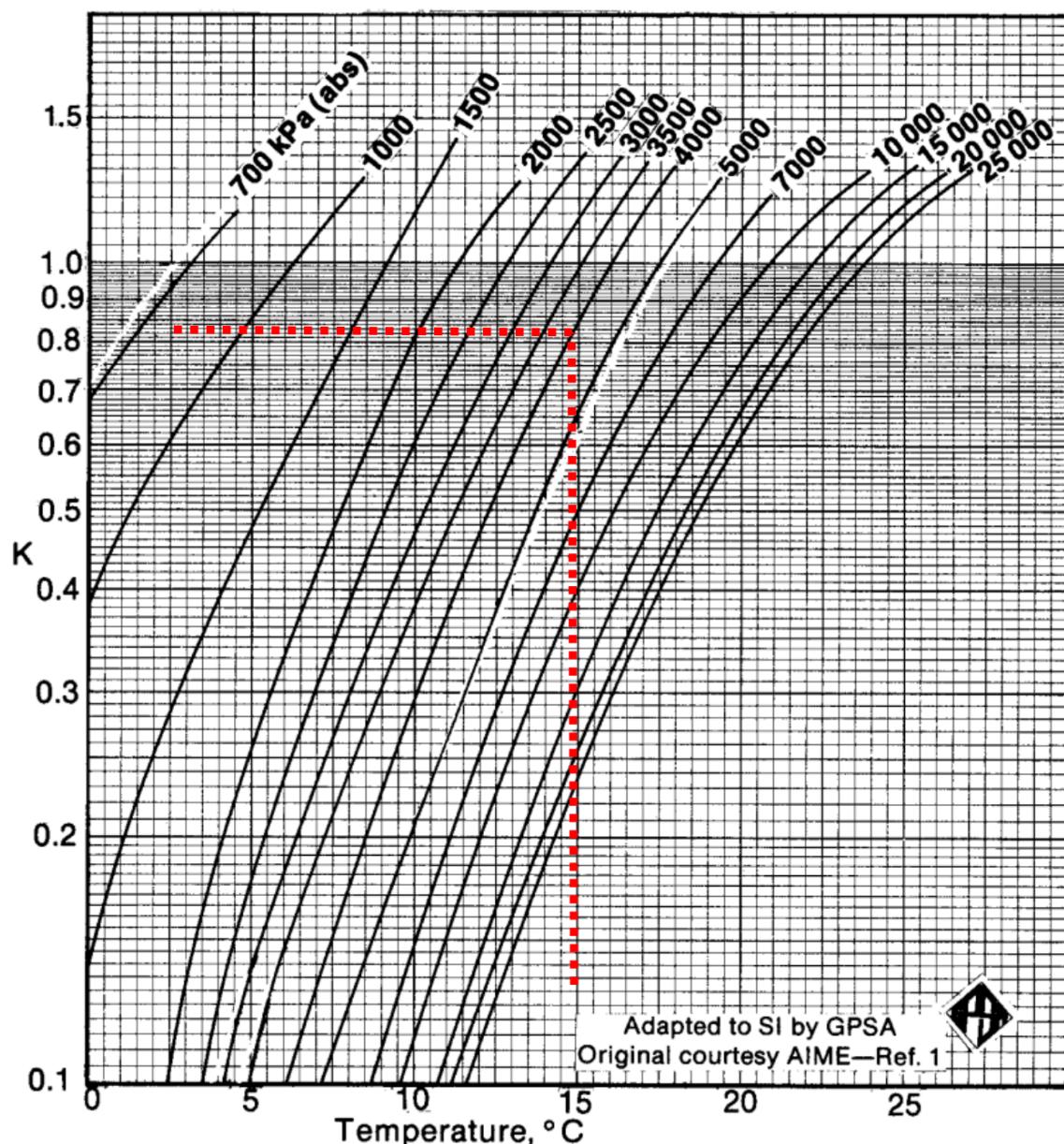
### Using the K-value chart

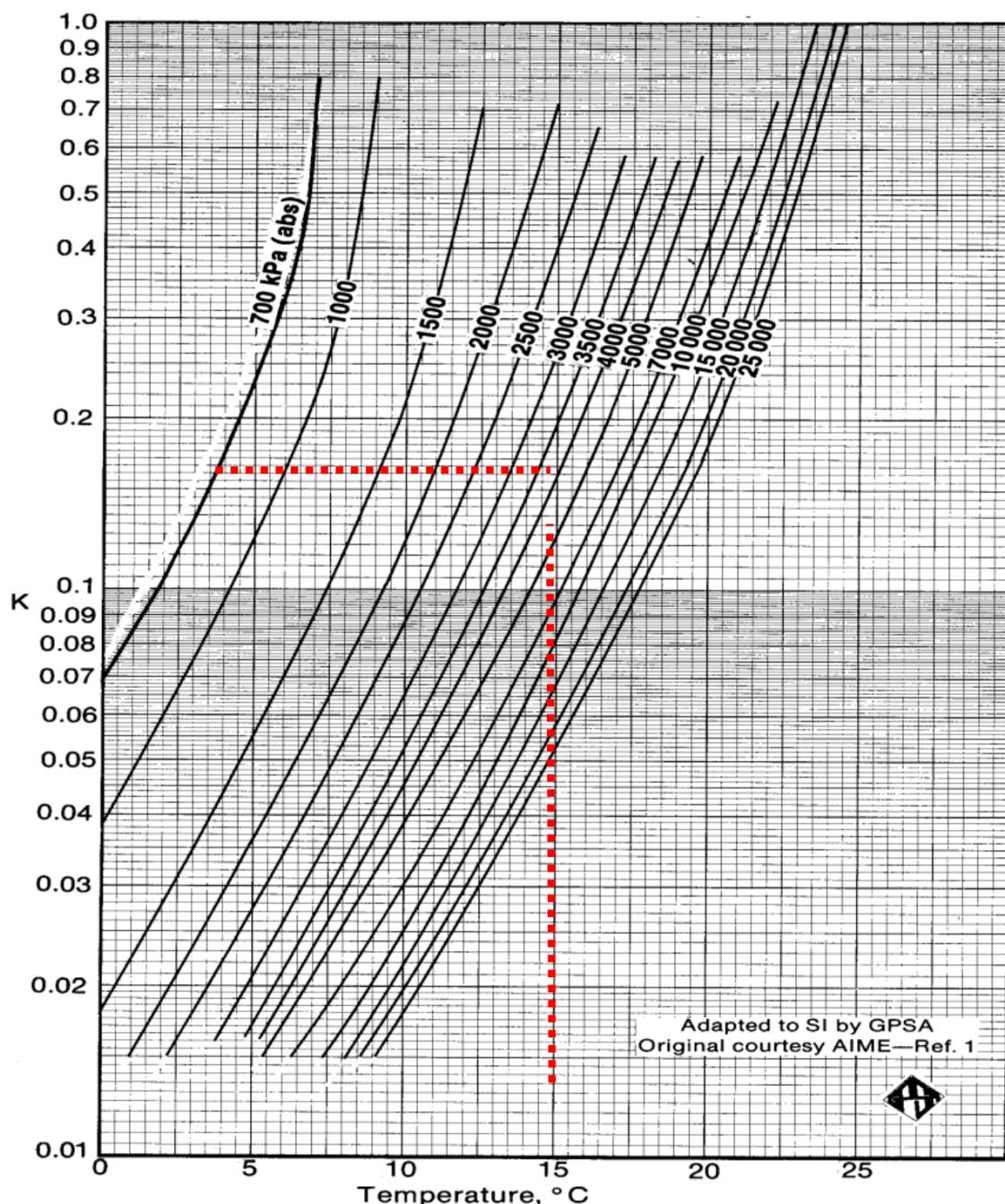
At the hydrate formation the mixture will satisfy

$$\sum_{i=1}^n \frac{y_i}{K_i} = 1 \quad (230)$$

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

$15\text{ }^{\circ}\text{C}$	$K_i$	$y_i$	$y_i/K_i$
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  $T = 17\text{ }^{\circ}\text{C}$  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.39

#### Question 39

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<sup>-1</sup>
- Inlet temperature  $T_1 = 70$  °C
- Sea temperature  $T_{\text{out}} = 4$  °C
- Fluid specific heat  $C_p = 1745$  J kg<sup>-1</sup> K<sup>-1</sup>

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

$$mC_p(T_1 - T_2) = UA\Delta T_{lm} \quad (231)$$

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

$$\Delta T_{lm} = \frac{(T_1 - T_{\text{out}}) - (T_2 - T_{\text{out}})}{\ln \left( \frac{T_1 - T_{\text{out}}}{T_2 - T_{\text{out}}} \right)} \quad (232)$$

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

$$U = \frac{mC_p}{A} \ln \left( \frac{T_1 - T_{\text{out}}}{T_2 - T_{\text{out}}} \right) \quad (233)$$

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

$$U = \frac{15 \frac{\text{kg}}{\text{s}} \cdot 1745 \frac{\text{J}}{\text{kg K}}}{\pi \cdot (0.292 \text{ m} + 2 \cdot 0.0127 \text{ m}) \cdot 15000 \text{ m}} \ln \left( \frac{70 \text{ °C} - 4 \text{ °C}}{18 \text{ °C} - 4 \text{ °C}} \right) \quad (234)$$

$$= 2.71 \frac{\text{W}}{\text{m}^2 \text{ K}} \quad (235)$$

**[Question end]**

**Q.40****Question 40**

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  $90 \text{ kg m}^{-3}$  and the liquid density is  $840 \text{ kg m}^{-3}$ . The initial temperature is  $70^\circ\text{C}$  and the surrounding sea temperature is  $4^\circ\text{C}$ . The hydrate forming temperature is  $15^\circ\text{C}$ .

**Solution:**

Rearrange the temperature decay equation

$$T = T_{\text{out}} + (T_{\text{initial}} - T_{\text{out}}) e^{\frac{UA}{mC_p} t} \quad (236)$$

to obtain an expression for the time

$$t = -\frac{mC_p}{UA} \ln \left( \frac{T - T_{\text{out}}}{T_{\text{initial}} - T_{\text{out}}} \right) \quad (237)$$

**liquid-filled section**

The total mass of liquid in the section is

$$m_L = \rho_L V = 840 \frac{\text{kg}}{\text{m}^3} \cdot \frac{\pi \cdot (0.17 \text{ m})^2 \cdot 10 \text{ m}}{4} \quad (238)$$

$$= 190.66 \text{ kg} \quad (239)$$

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

$$t = -\frac{190.66 \text{ kg} \cdot 3100 \frac{\text{J}}{\text{kg K}}}{1.5 \frac{\text{W}}{\text{m}^2 \text{K}} \cdot (\pi \cdot 0.2 \text{ m} \cdot 10 \text{ m})} \ln \left( \frac{15^\circ\text{C} - 4^\circ\text{C}}{70^\circ\text{C} - 4^\circ\text{C}} \right) \quad (240)$$

$$= 112365 \text{ s} = 31.2 \text{ hrs} \quad (241)$$

**gas-filled section**

The total mass of gas in the section is

$$m_G = \rho_G V = 90 \frac{\text{kg}}{\text{m}^3} \cdot \frac{\pi \cdot (0.17 \text{ m})^2 \cdot 10 \text{ m}}{4} \quad (242)$$

$$= 20.43 \text{ kg} \quad (243)$$

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

$$t = -\frac{20.43 \text{ kg} \cdot 3100 \frac{\text{J}}{\text{kg K}}}{1.5 \frac{\text{W}}{\text{m}^2 \text{K}} \cdot (\pi \cdot 0.2 \text{ m} \cdot 10 \text{ m})} \ln \left( \frac{15^\circ\text{C} - 4^\circ\text{C}}{70^\circ\text{C} - 4^\circ\text{C}} \right) \quad (244)$$

$$= 12040 \text{ s} = 3.3 \text{ hrs} \quad (245)$$

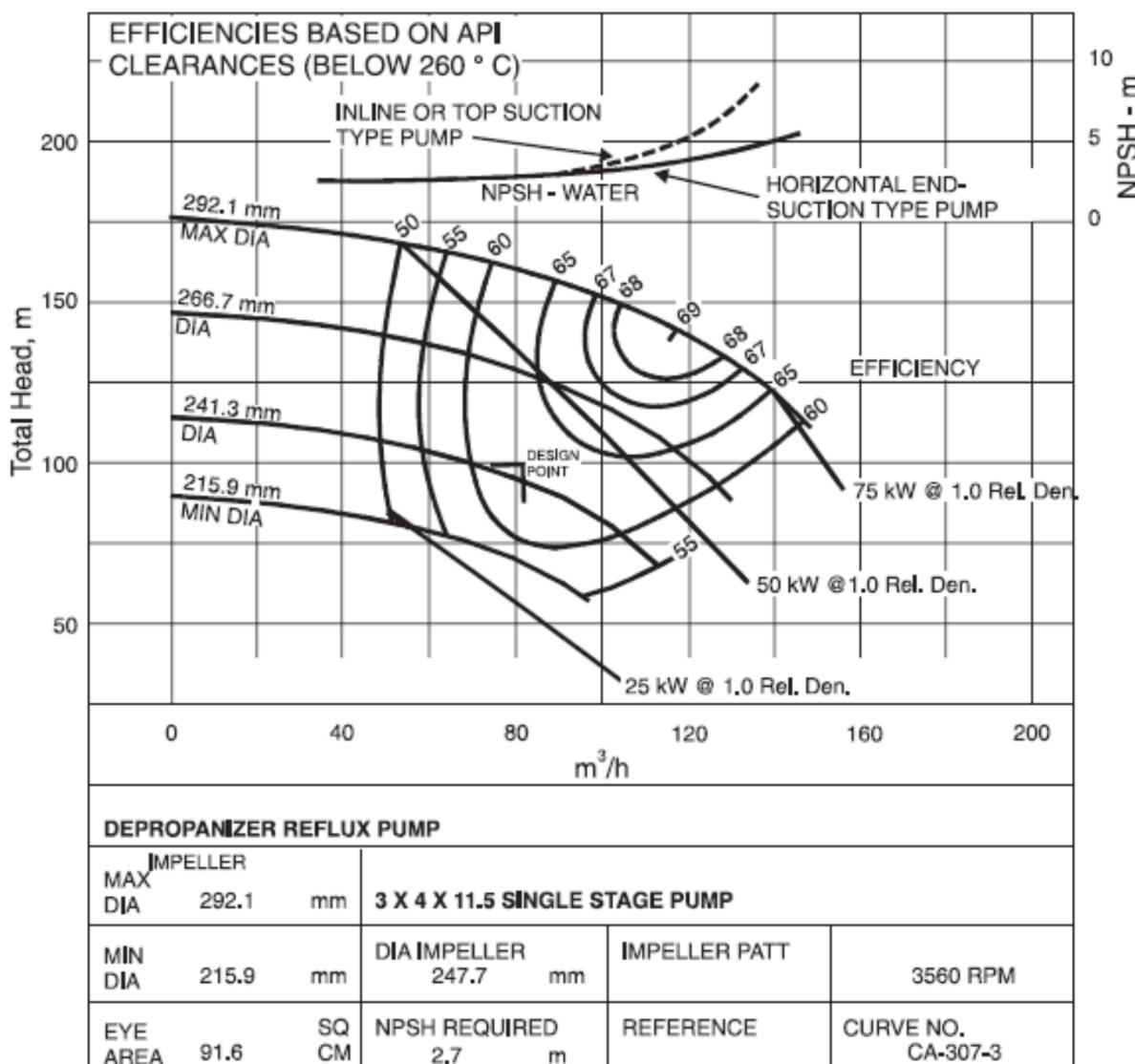
**[Question end]**

## 9 Energy efficiency

Q.41

### Question 41

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

$$\frac{q_1}{q_2} = \left( \frac{N_1}{N_2} \right) \cdot \left( \frac{d_1}{d_2} \right)^2 \quad (246)$$

$$\frac{H_1}{H_2} = \left( \frac{N_1}{N_2} \right)^2 \cdot \left( \frac{d_1}{d_2} \right)^2 \quad (247)$$

$$\frac{W_1}{W_2} = \left( \frac{N_1}{N_2} \right)^3 \cdot \left( \frac{d_1}{d_2} \right)^3 \quad (248)$$

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

$$H_1 = 73 \text{ m} \cdot \left( \frac{266.7 \text{ mm}}{215.9 \text{ mm}} \right)^2 = 111.4 \text{ m} \quad (249)$$

$$W_1 = 27 \text{ kW} \cdot \left( \frac{266.7 \text{ mm}}{215.9 \text{ mm}} \right)^3 = 50.9 \text{ kW} \quad (250)$$

**[Question end]**

#### Q.42

#### Question 42

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

$$NPSH_{\text{available}} = \frac{10.19 (p_s - p_v - \Delta p_f)}{\gamma} + H_s - H_m - \frac{v^2}{2g} - H_{ac} \quad (251)$$

where  $p_s$  is the pressure inside the suction vessel,  $p_v$  is the vapour pressure of the liquid at pump suction,  $\Delta p_f$  is the friction loss in the suction line,  $\gamma$  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

$$NPSH_{\text{available}} = \frac{10.19 (-\Delta p_f)}{\gamma} + H_s \quad (252)$$

$$= \frac{10.19 \cdot -0.075 \text{ bar}}{0.502} + (5 \text{ m} - 1 \text{ m}) \quad (253)$$

$$= 2.48 \text{ m} \quad (254)$$

- *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.*

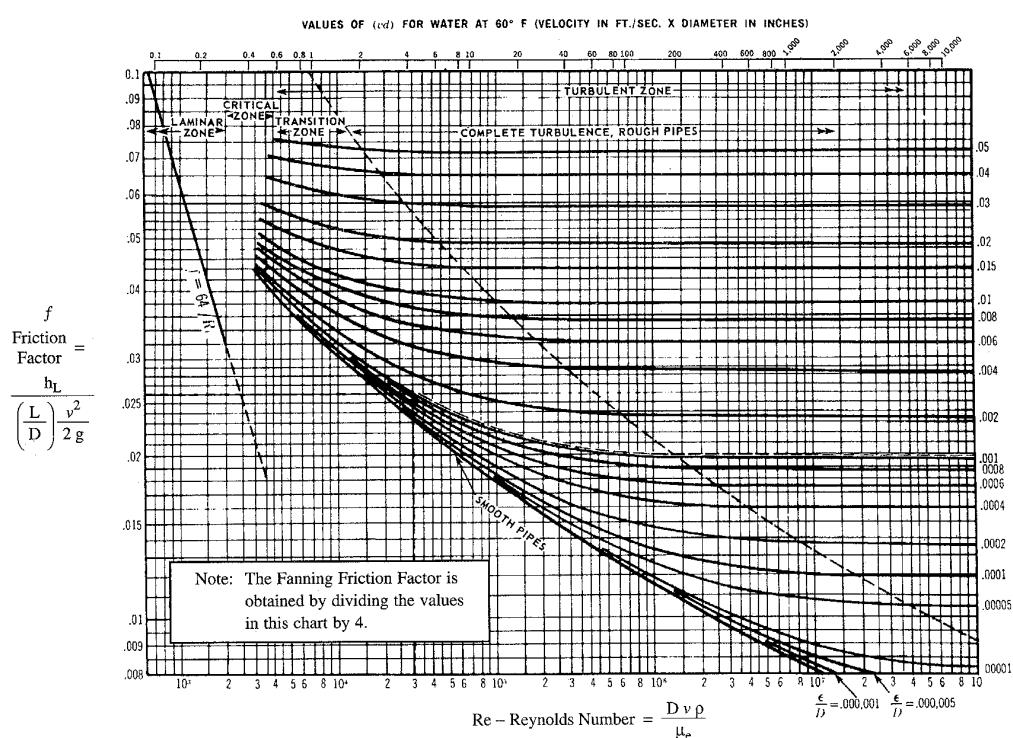
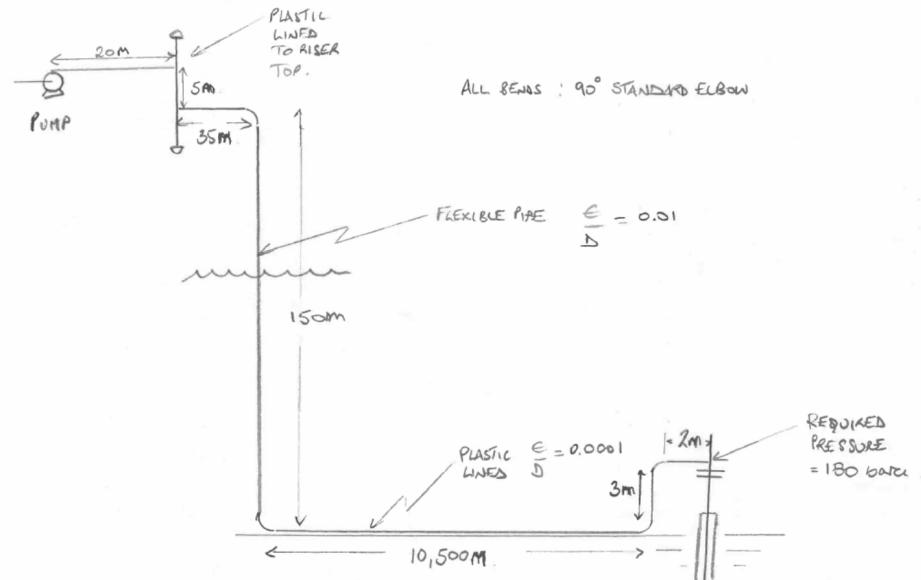
Adding a cooler in the pump suction line will reduce the vapour pressure while the suction pressure remains constant resulting in a greater available NPSH.

### [Question end]

#### Q.43

#### Question 43

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<sup>-3</sup> 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

**Solution:****[Question end]****Q.44 Question 44**

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.45**

**Question 45**

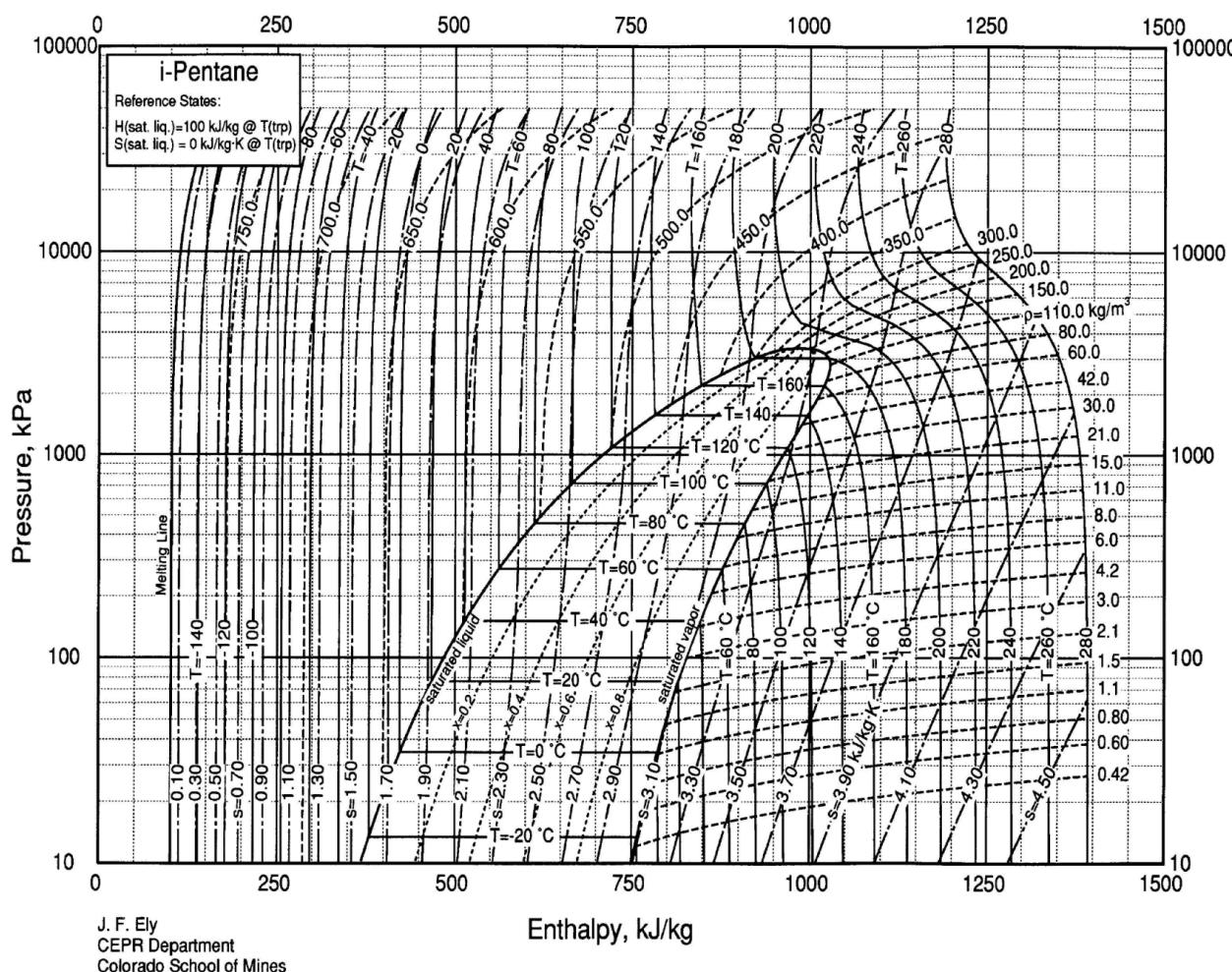
A high pressure gas condensate well flows at  $30,000 \text{ kg hr}^{-1}$ . The arrival pressure and temperature are 650 bara and  $160^\circ\text{C}$ . The pressure is dropped across a choke valve to 100 bara. If the choke were replaced by an expander how much energy could be produced? The inlet and 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.46**

**Question 46**

i-pentane is to be utilised in an Organic Rankine Cycle (ORC). 100,000 BPD of hot water is available at  $100^\circ\text{C}$ . Cooling water is available at  $20^\circ\text{C}$ . Use the attached Mollier chart and estimate the potential for power recovery in kW.



[Question end]

**Q.47**

**Question 47**

$15 \text{ kg s}^{-1}$  of methane is to be compressed from 5 bara to 15 bara. The suction temperature to the compressor is  $30^\circ\text{C}$ .

- 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.48 Question 48

A reciprocating compressor has 6 cylinders each with a total cylinder volume of 0.002 m<sup>3</sup>. 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.49 Question 49

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

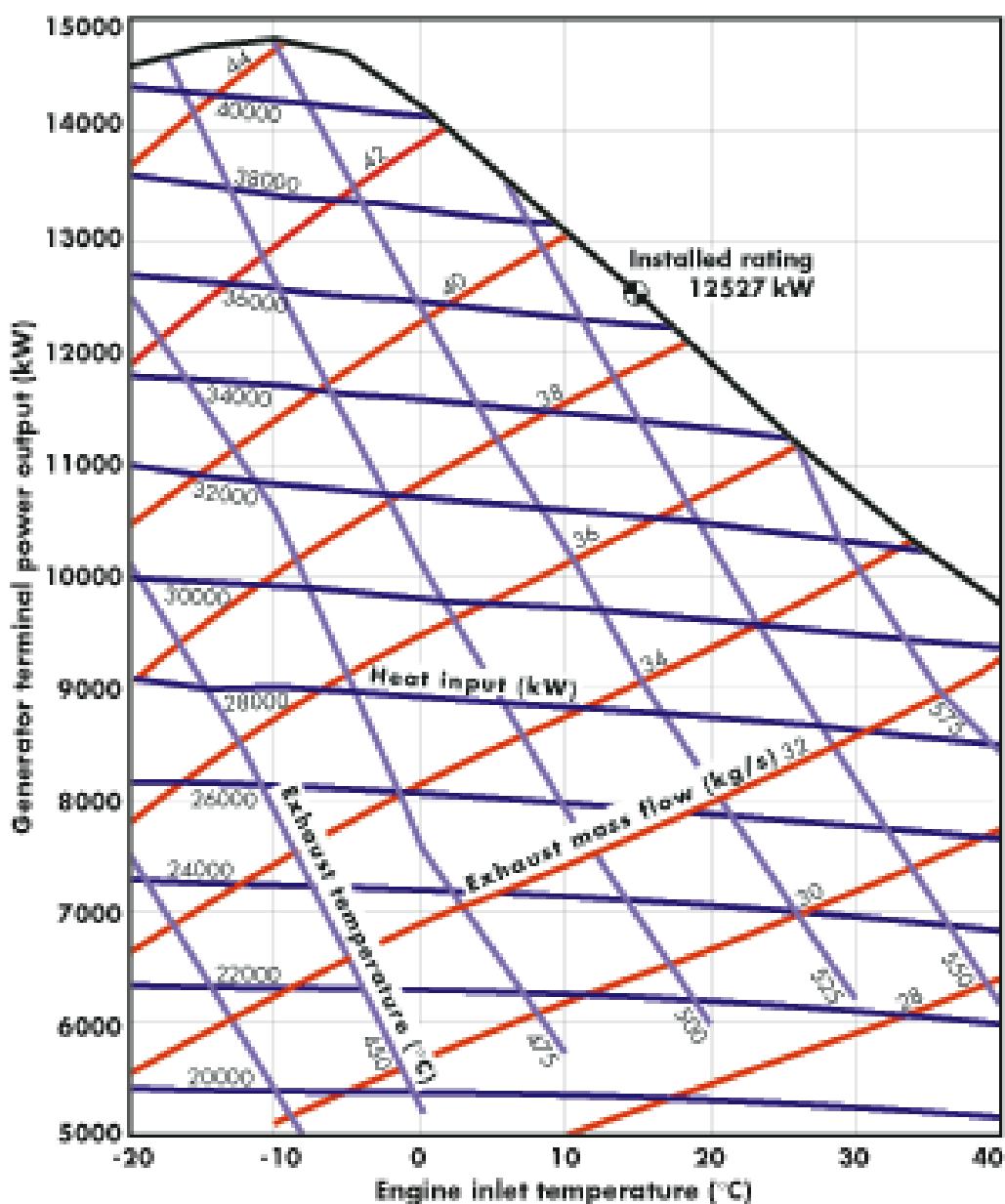
By conducting an annual review of machine efficiencies, determine which GT would be the most efficient over the field life.

[Question end]

#### Q.50 Question 50

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]

## 10 Water injection

### Q.51 Question 51

Calculate the dissolved oxygen concentration of seawater at 5 ° C and 20 ° C.

[Question end]

### Q.52 Question 52

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  $8 \times 10^{-6}$  mol fraction, the

seawater leaves the bottom of the column with a concentration of  $0.05 \times 10^{-6}$  mol fraction. The column operates at  $10^\circ\text{C}$ . 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  $0.04 \times 10^{-6}$  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 $^\circ\text{C}$	$H_{O_2}$ 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]