

6. Gas Treatment

Gas Sales Specifications

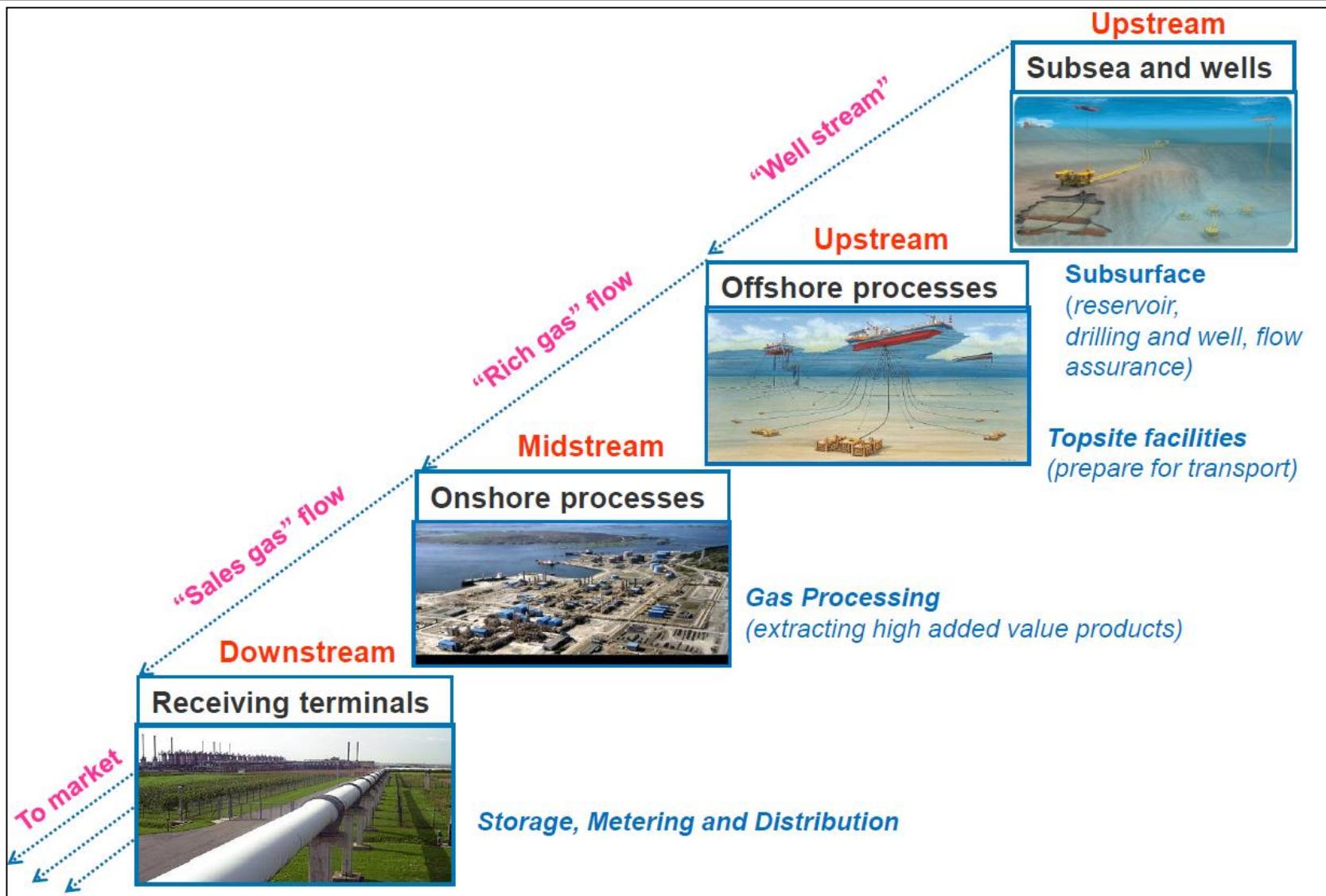
In order to condition gas for sales distribution a range of quality specifications require to be achieved. These vary from country to country, common features are hydrocarbon and water dewpoint, temperature, pressure and composition e.g. H₂S, CO₂, mercury and sulphur compounds

Water dewpoint limits are required to avoid corrosion and hydrate formation, and depend on typical ambient conditions. May be stated as a dewpoint (e.g. -10 deg C at 69 barg) or as a water content (e.g 2lb/ mmscf, 30mg/Sm³).

Temperature - A maximum temperature at the delivery point may be specified, usually around 30-50 °C

Pressure - The maximum gas pressure will be decided by the design pressure of the system and the allowable back pressure on other system entrants. Nominal gas pressure is the normal entry pressure to the pipeline, typically 70 - 140 Bar

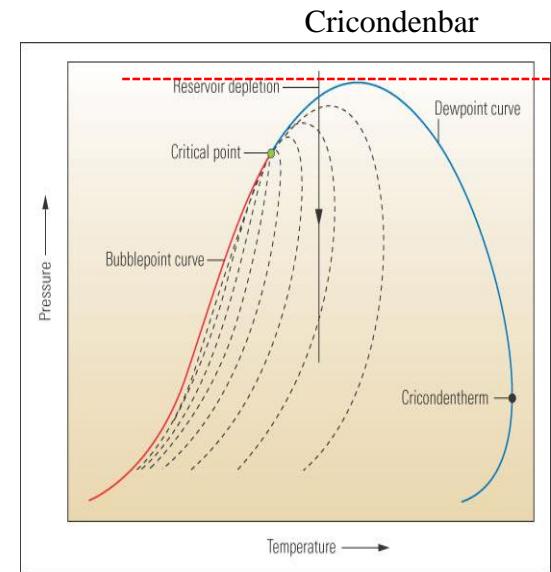
Gas Value Chain



Gas Sales

Hydrocarbon Dewpoint Control

- To prevent hydrocarbon condensation, with the consequences of the pipeline flowing two-phase, the pipeline operator often sets a limit on the gas cricondenbar. The significance of the cricondenbar is evident from the phase envelope - provided system pressures are higher than the cricondenbar then a single phase will always exist irrespective of temperature. A typical cricondenbar specification is 105-110 bara maximum.
- An alternative to cricondenbar control is a hydrocarbon dewpoint or a liquid loading maximum value may be given



Solids: Free of particulates in amounts detrimental to transmission and utilisation equipment

Gas Sales

The sales gas specification will be subject to a pricing agreement which is likely to include the following:

Gross Calorific Value (GCV) or Higher Heating Value (HHV)

The total heat produced by combustion of the fuel

Net Calorific Value (NCV) or Lower Heating Value (LHV)

The total heat produced by combustion of the fuel minus the latent heat contained in the water vapour discharged as fuel gas,
NCV represents the available heat

Wobbe Index (WI)

Wobbe Index is used to compare fuel quality for different gases
it characterises flame stability

WI is the ratio of GCV to the square root of the gas s.g. - units are MJ/Sm³

$$WI = GCV / (\text{Gas sg})^{0.5}$$

Sulphur Content

Sulphur content is controlled for safety reasons, to prevent pipeline corrosion and improve the sales value of the gas

Gas	GCV of dry gas at 15 °C & 101.325 kPa (MJ/m ³)
Methane	37.69
Ethane	66.03
Propane	93.97
i-Butane	121.43
n-Butane	121.78
i-Pentane	149.32
n-Pentane	149.65
Hexane	177.56
Heptane	205.43
CO ₂	0
H ₂ S	23.79

Typical Gas Pipeline Specification

Gas Cricondenbar < 106 bara

As calculated by the Peng Robinson equation of state or such alternative method as may be agreed from time to time.

Pressure 110 to 149.9 bara

Temperature -5° C to 55° C

Water Content < 24 kg/Sm³

H₂S Content < 2.3 ppm (based on present equipment)

Total Sulphur Content < 15 ppm expressed as H₂S

CO₂ Content < 3.8% mol

O₂ Content < 7.5 ppm

Mercury Maximum mercury content of 0.01 micrograms per Cubic Metre, unless otherwise agreed

Mercaptans < 1.0ppm v/v

Typical Gas Pipeline Specification



Gas shall be commercially free from objectionable odours and from materials and dust or other solid or fluid matter, waxes, gums and gum forming constituents which might cause injury to, or interference with, the proper operation of the lines, meters, regulators or other appliances or facilities through which it flows.

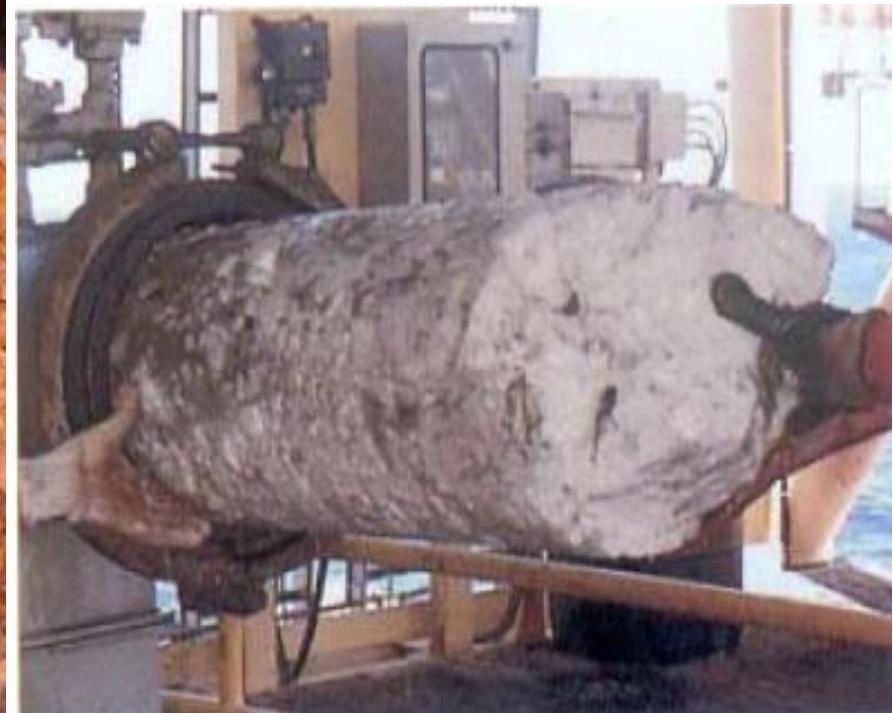
No chemicals or other substances may be introduced by any means into the Gas, which might be carried over into the transportation system without the prior agreement of the pipeline operator, such agreement not to be unreasonably withheld.

Shipper Gas must be capable of being processed to meet National Grid redelivery specification on Wobbe Index, Incomplete Combustion Factor, and Soot Index.

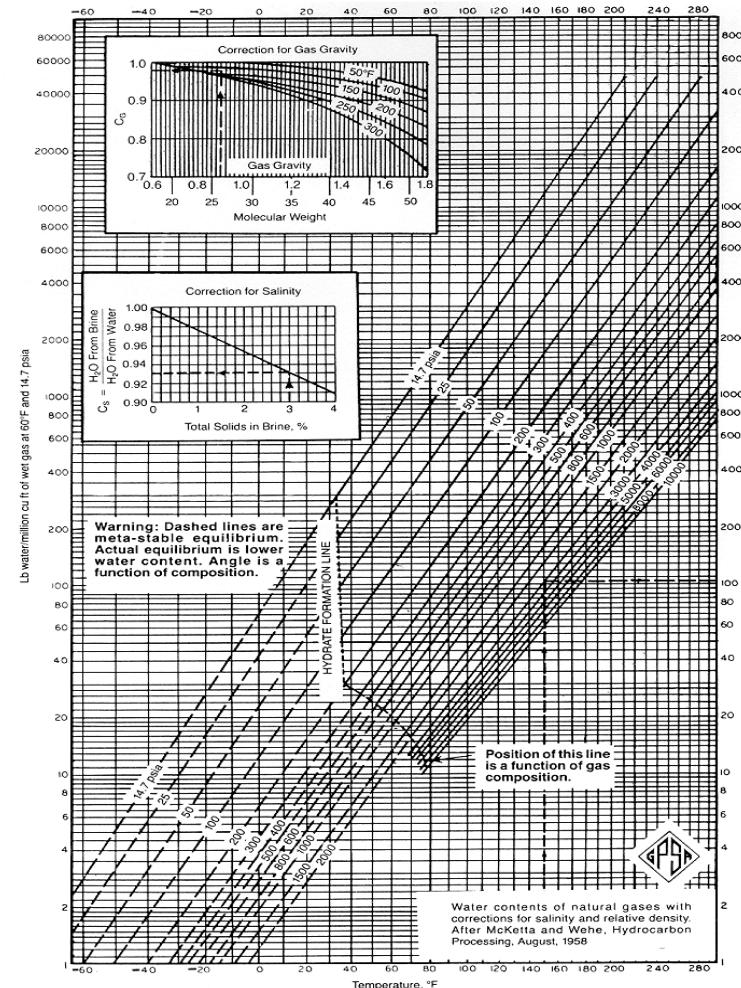
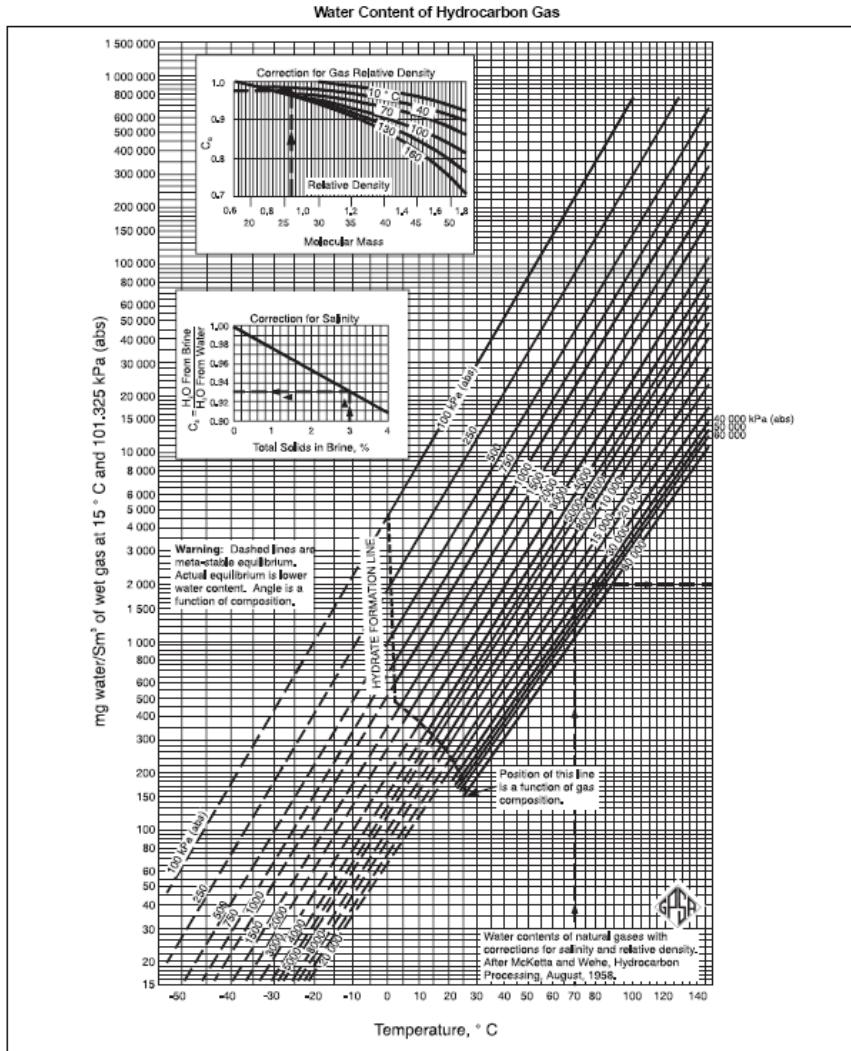
The above specification may be subject to change in the future as a result of different requirements from regulators, or governmental authorities, not within the control of the pipeline operator.

The gas will be redelivered from the Terminal to the National Grid Terminal for entry into the National Transmission System (NTS) in a condition to meet the NTS Specification, with the CO₂ content in accordance with the specification applicable to the Frigg UK Terminal (currently 4% mol). Depending on the CO₂ content of a new entrant FUKA may be able to facilitate a CO₂ blending service.

Water – Corrosion and Hydrates



Water Content of Natural Gas



Water Content of Natural Gas

Gas water content is a function pressure, temperature and gas composition.

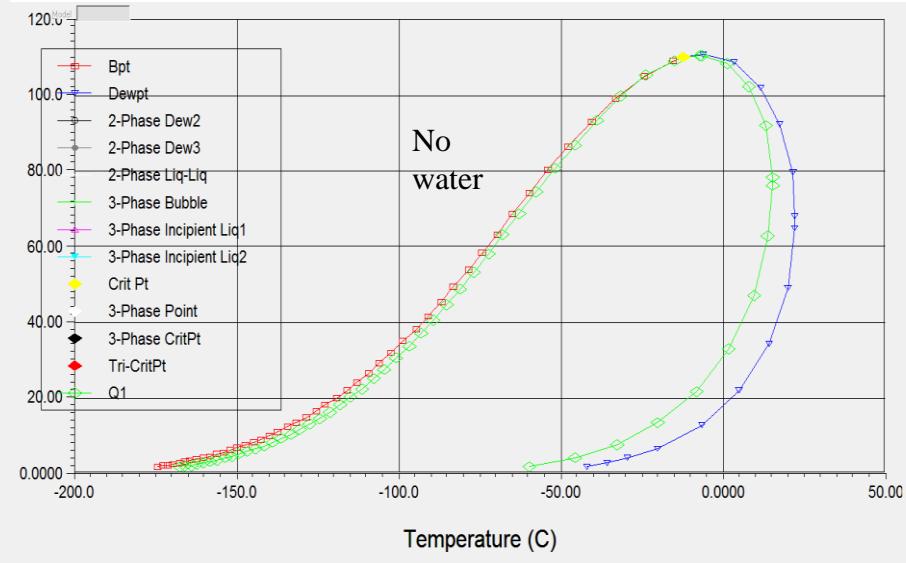
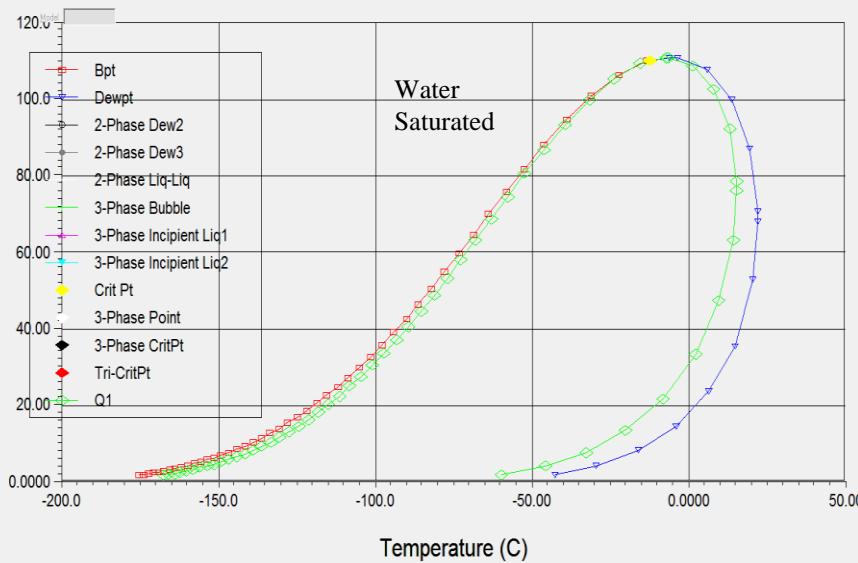
Methods for determining water include;

Vapour/partial pressure

Empirical charts

Equations of State

Effect of Water on Phase Envelope



Critical Values

Two-Phase Critical Temperature	-12.23 C
Two-Phase Critical Pressure	110.3 bar
Three-Phase Critical Temperature	<empty>
Three-Phase Critical Pressure	<empty>

Maxima

Cricondentherm	22.33 C
Cricondenbar	111.0 bar

Critical Values

Two-Phase Critical Temperature	-12.22 C
Two-Phase Critical Pressure	110.1 bar
Three-Phase Critical Temperature	<empty>
Three-Phase Critical Pressure	<empty>

Maxima

Cricondentherm	22.34 C
Cricondenbar	110.8 bar

Hydrate Formation

Presence of water in gas transportation system may result in hydrate formation and/or corrosion

Hydrate Formation

- presence of free water

- low temperature

- high pressure

- C_1-C_4 paraffins

- H_2S/CO_2

Corrosion

- partial pressure H_2S/CO_2

- temperature

- water pH

Hydrate formation boundary estimated by;

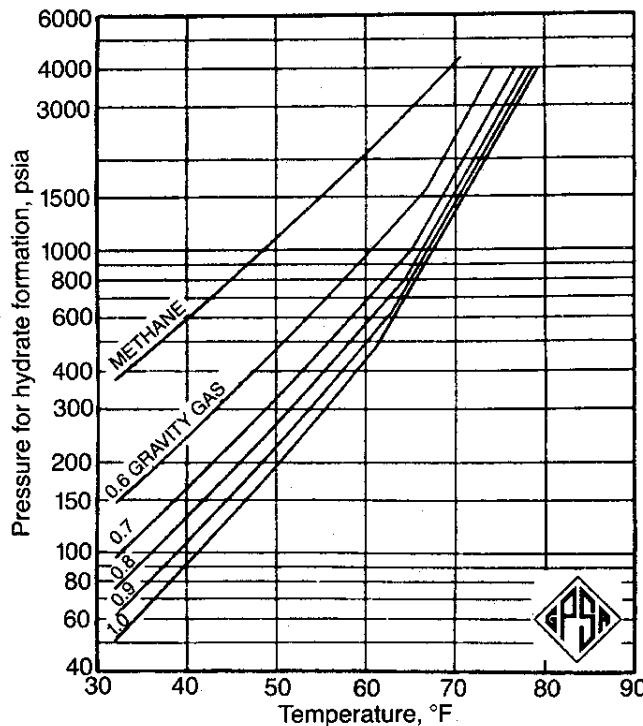
- Empirical charts (on right)

- K (equilibrium) factors

- equations of state

- modules within most commercial simulators

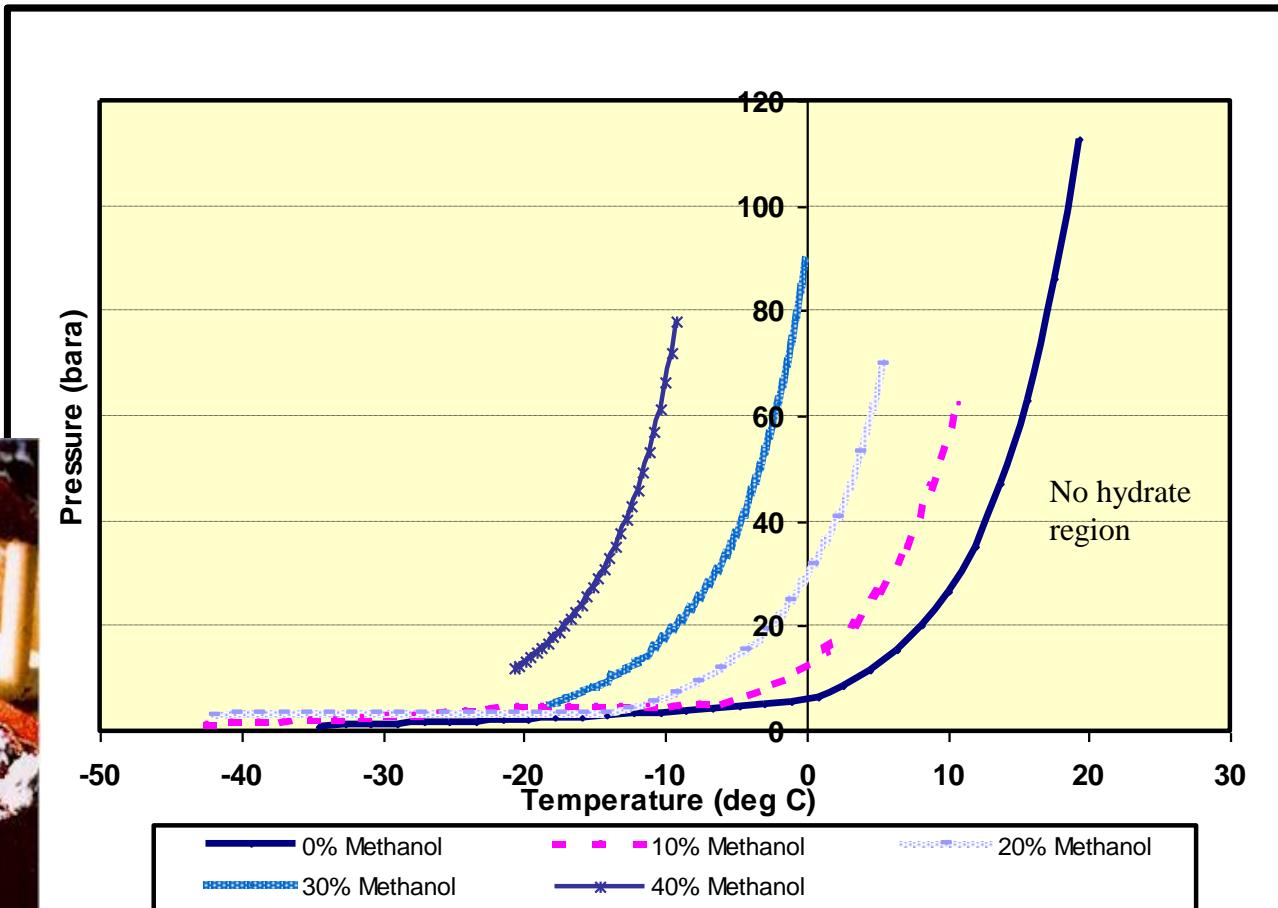
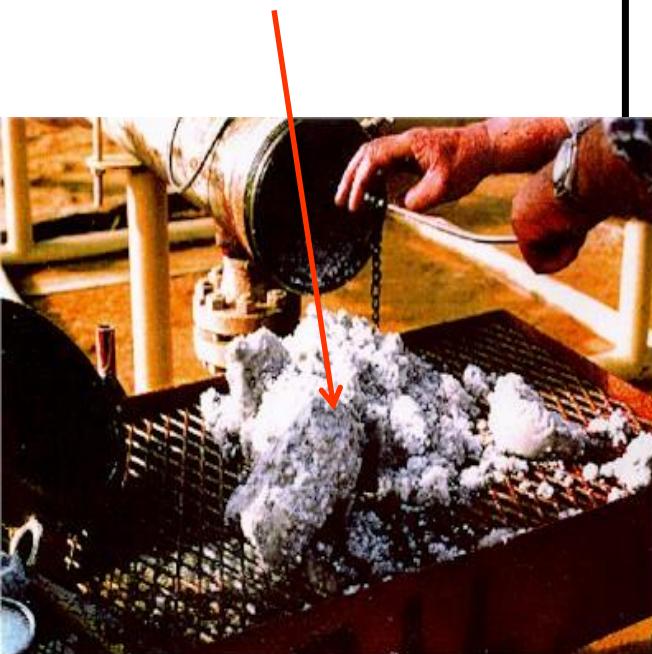
Pressure-Temperature Curves For Predicting Hydrate Formation



Note that this figure should only be used for first approximations of hydrate formation conditions

Hydrate Curves

At atmospheric pressure hydrate will be unstable and subliming to gas and water.



Hydrate Inhibition

Inhibition or hydrate formation boundary suppression can be facilitated by injecting;

- methanol
- glycol
- threshold hydrate inhibitors

The Hammerschmidt equation is often used for prediction of the necessary inhibitor concentration to achieve a given amount of hydrate depression:

$$X = 100 \cdot \frac{d \cdot M_i}{K_i + d \cdot M_i}$$

X Weight percent of inhibitor in the liquid water phase (wt%)

d Depression of hydrate point ($^{\circ}\text{C}$) – hydrate formation temperature less minimum pipeline temperature.

M_i Mol weight of inhibitor (MW methanol = 32)

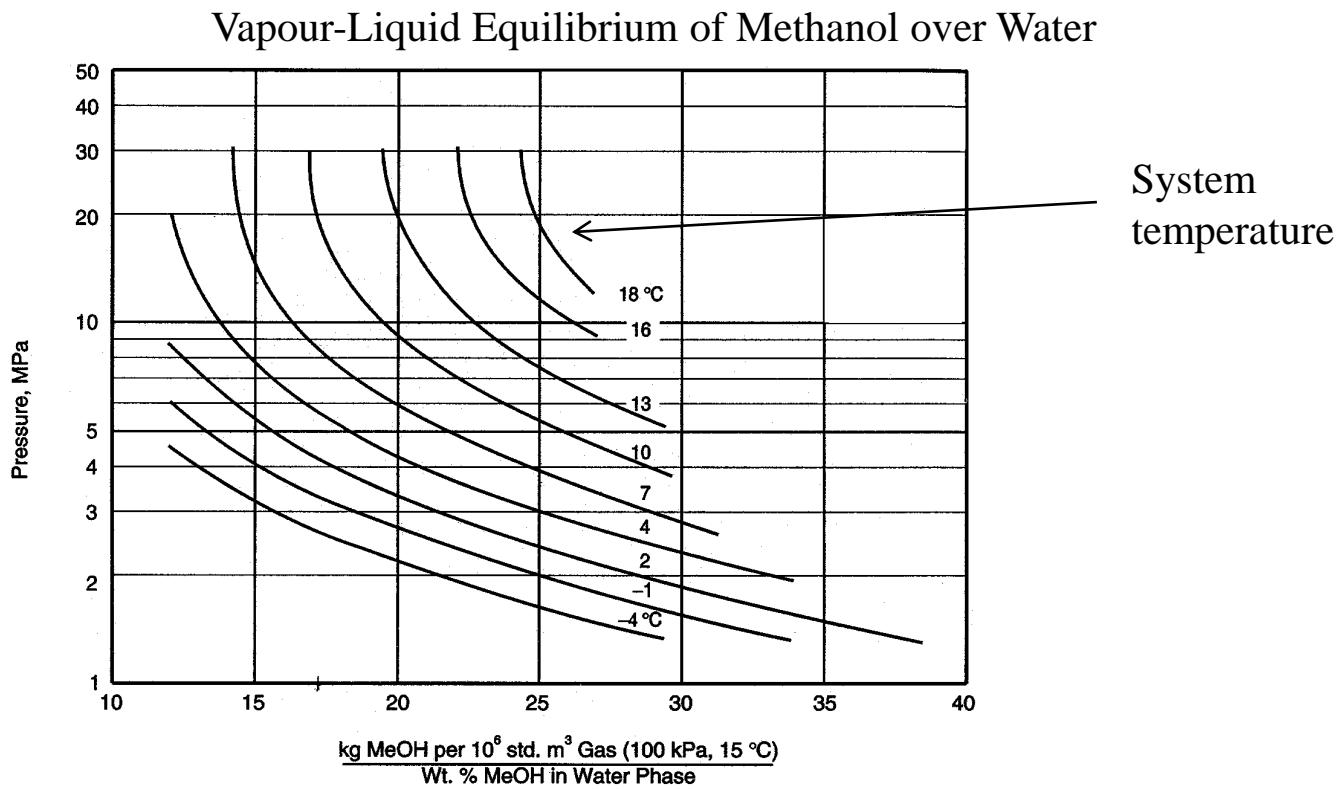
K_i Constant:

1297 for methanol

2220 for glycols

Hydrate Inhibition

When methanol or glycol is added for hydrate prevention some of the methanol or glycol will vapourise. The amount of hydrate inhibitor which vapourises must be taken into account when estimating inhibitor injection rates. The following chart can be used for estimating methanol losses to the vapour phase.



Inhibition Worked Example

10 MMscfd ($2.83 \times 10^5 \text{ stm}^3/\text{d}$) of natural gas (specific gravity = 0.65) cools to 4.3°C in a subsea pipeline. The pipeline pressure is 6.2 MPa, outlet is 5.5 MPa. What weight % methanol ($\rho = 800 \text{ kg/m}^3$, Molecular Weight = 32) must be added to prevent hydrate formation if the gas enters the line water saturated at 32.2°C and what is the rate of injection ?

(FOLLOWING SLIDES ILLUSTRATE INTERMEDIATE STEPS WITH GRAPHS)

Calculate the hydrate formation point from GPSA Chart = 21 DegC (16 plus 5 DegC safety margin)

Calculate required hydrate point depression:

$$d = 21.0 - 4.3 = 16.7^\circ\text{C}$$

Use GPSA chart for water content of lean natural gas and calculate the amount of liquid water formed in pipeline:

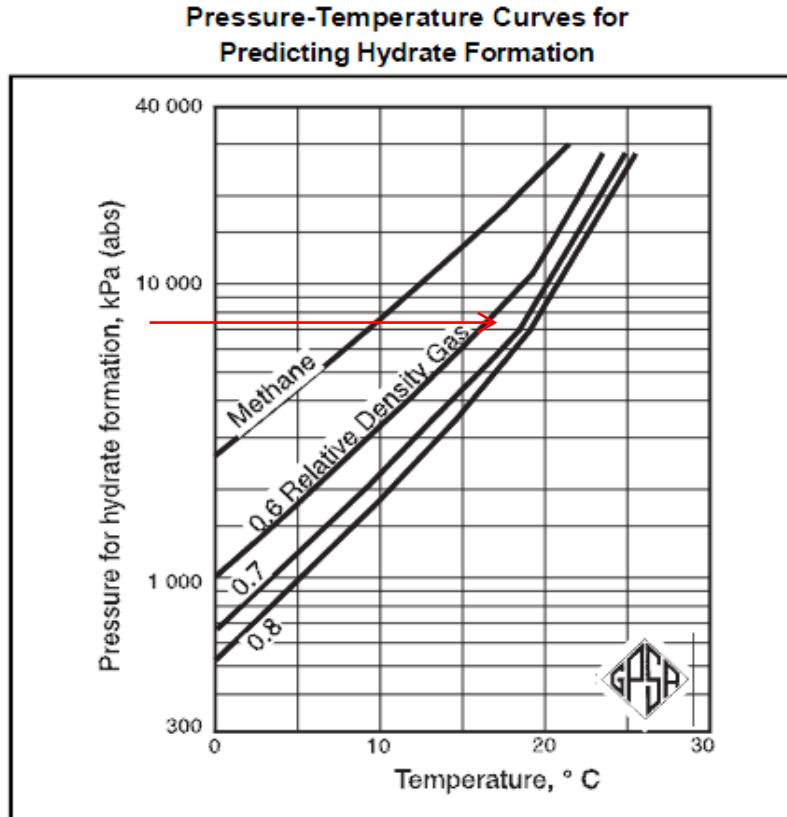
Water content @ 32.2°C and 6.4 Mpa = $800 \text{ kg}/10^6 \text{ stm}^3$

Water content @ 4.4°C and 5.5 Mpa = $\underline{220 \text{ kg}/10^6 \text{ stm}^3}$

The difference is the amount of condensed water $580 \text{ kg}/10^6 \text{ stm}^3$

$$\text{Total liquid } H_2O = \left| \frac{580 \text{ kg}}{10^6 \text{ stm}^3} \right| \cdot \left| \frac{0.283 \times 10^6 \text{ stm}^3}{\text{day}} \right| = 164.1 \text{ kg/day}$$

Hydrate Formation Temperature



At 6.4 Mpa hydrate point is 16 Deg C.
 Add 5 Deg C safety margin. Use Hydrate formation temperature as 21 Deg C

Inhibition Worked Example

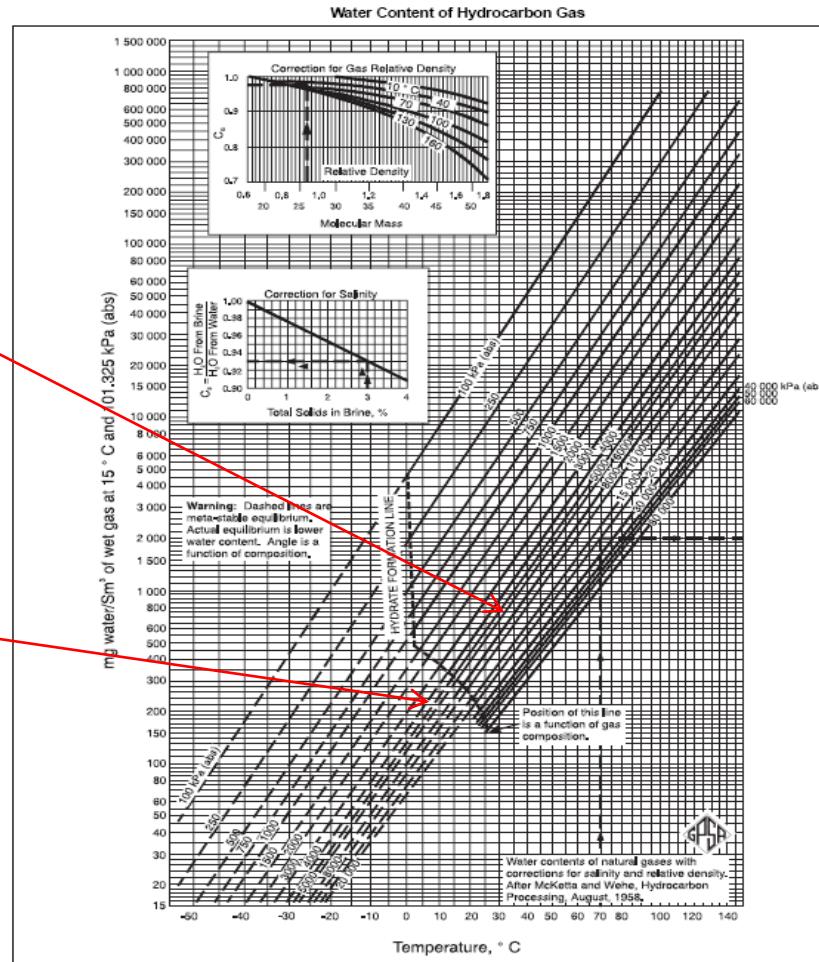
32.2 Deg C and 6.2Mpa

Water Saturation = 800 mg water/Sm³ (= kg/millionSm³)

(ignore correction)

4.3 Deg C and 5.5Mpa

Water Saturation = 220 mg water/Sm³



Inhibition Worked Example

Use Hammerschmidt equation to calculate methanol concentration :

$$X = 100 \cdot \frac{d \cdot M_i}{K_i + d \cdot M_i}$$

$$X_{MeOH} = 100 \cdot \frac{(16.7)(32)}{1297 + (16.7)(32)} = 29.2 \text{ wt\%}$$

$K_i = 1297$ for methanol

Calculate the required mass of methanol in water:

$$m_{MeOH} = 164.1 \cdot \left[\frac{29.2}{100 - 29.2} \right] = 67.7 \text{ kg / day}$$

Calculate the methanol loss to the vapour phase using the vapour-liquid equilibrium chart (4.4°C, 5.5 MPa):

$$\text{Vaporisation losses} = 17 \cdot \frac{\text{kg MeOH} / 10^6 \text{ stm}^3}{\text{wt\% MeOH in water phase}}$$

$$m_{MeOH} = \frac{17 \text{ kg MeOH}}{10^6 \text{ stm}^3} \cdot \frac{0.283 * 10^6 \text{ stm}^3}{\text{day}} \cdot 29.2 \text{ wt\% MeOH}$$

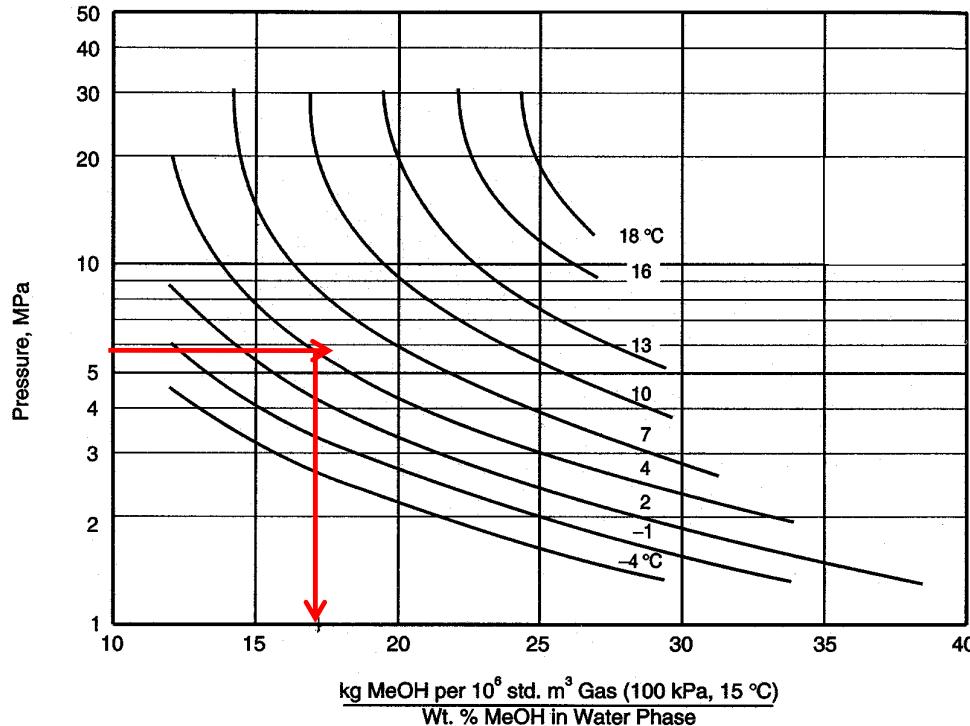
$$= 140 \text{ kg/day}$$

Total injection rate = $67.7 + 140 = 207.7 \text{ kg/day}$

or

$207.7 / 0.8 = 260$ litres/day of methanol requires to be injected into pipeline

Inhibition Worked Example



Gas Dehydration

- Glycol/Methanol Injection
- Glycol (TEG) Contacting
 - Basic
 - Cold Finger
 - OPC DRIZO
- Solid Dessicants
 - Molecular Sieves
 - Silica Gels
- Other Technology
 - Membranes
 - IFPEX
 - Shell twister

Comparison of Dehydration Methods

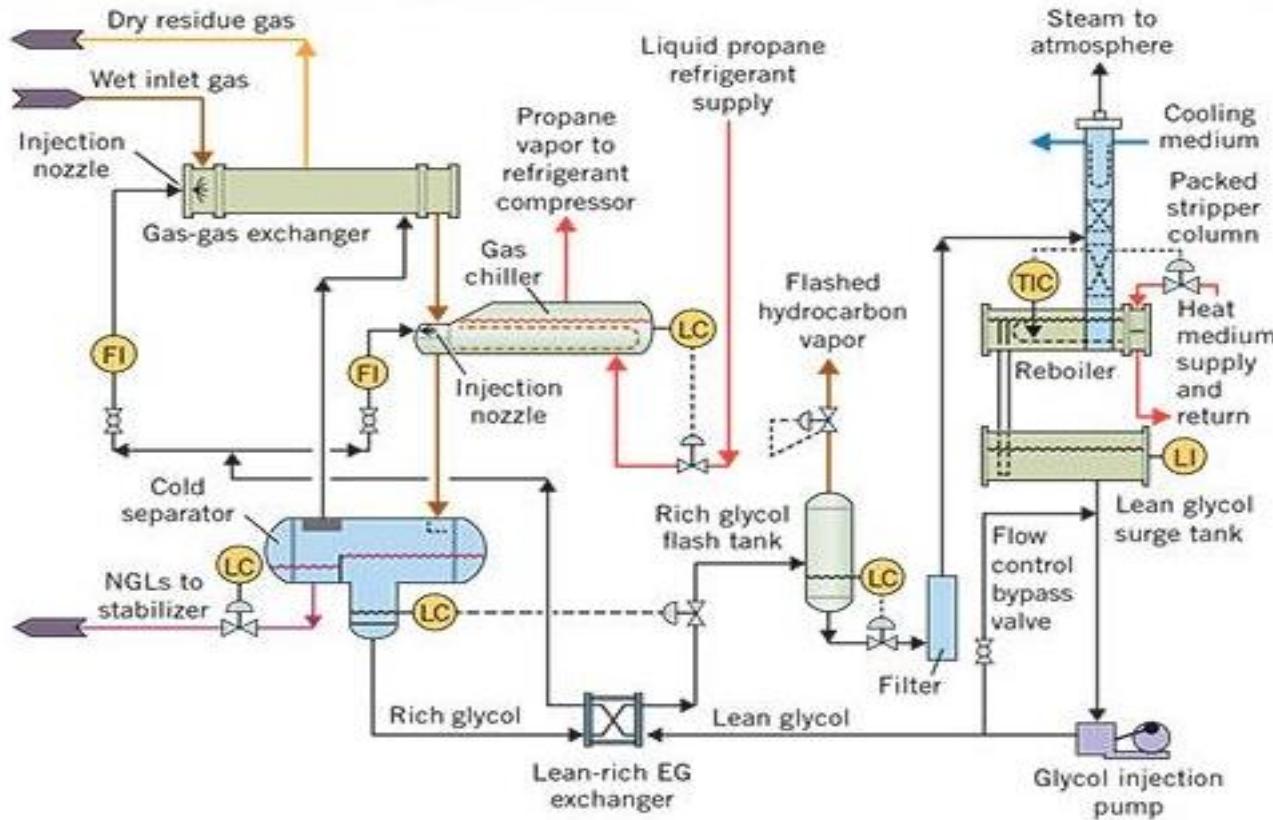
Method	Gas Specification (ppm v/v)	Gas Specification (lb/MMscf)	Remarks
TEG Contacting	20	1	Atmospheric regeneration Min. contact temp 15-26°C
	5	0.25	Vacuum regeneration and gas stripping
	2	0.1	Regeneration by azeotropic distillation (OPC Drizo)
Solid Desiccants	1	0.05	Molecular sieves
	2	0.1	Alumina
	10	0.5	Silica gel (high capital/operating costs)
Membranes	20	1	"Permea"

When to choose MEG or TEG dehydration?

<http://www.jmcampbell.com/tip-of-the-month/2007/03/meg-injection-vs-teg-dehydration/>

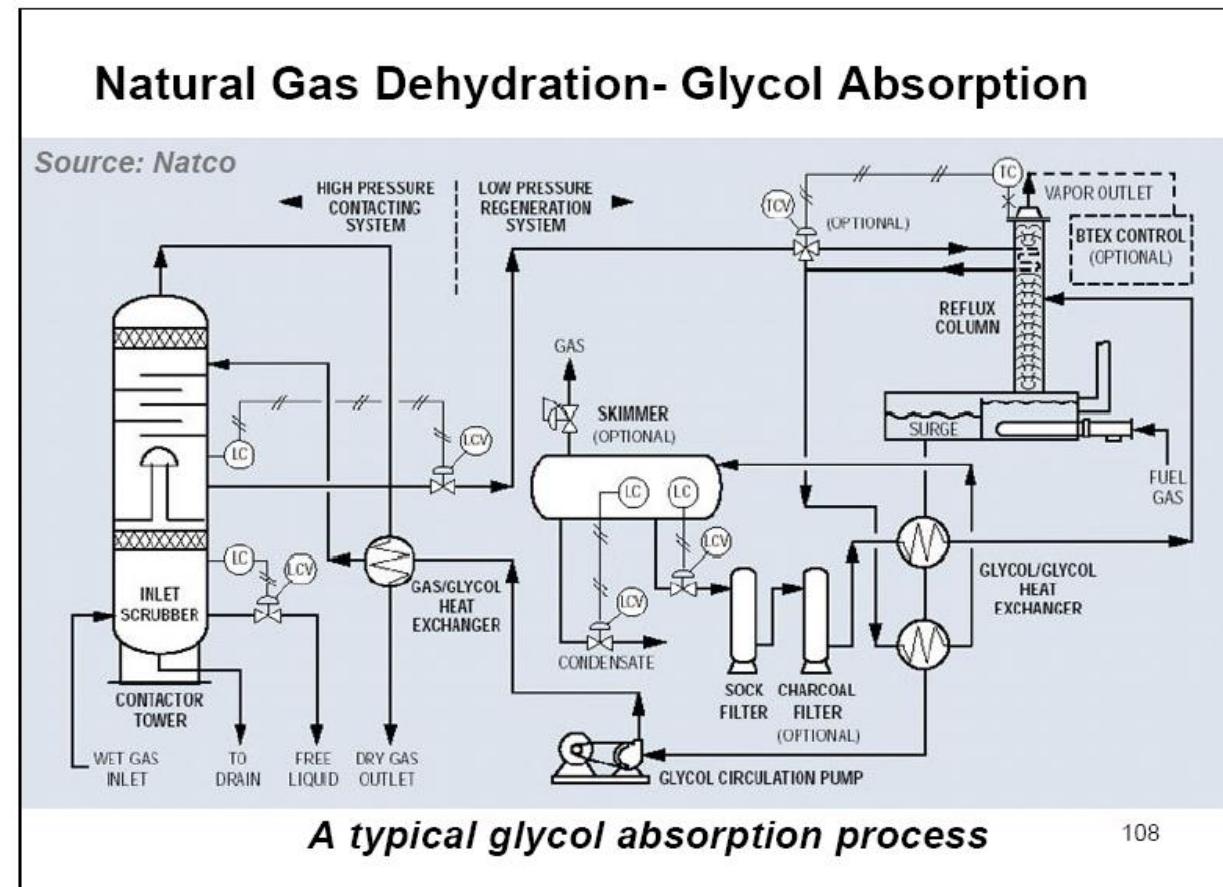
Glycol Injection

Exchanger and chiller (used to recover NGLs and remove water) at risk of hydrate formation -> Use MEG to prevent hydrates (not dehydrating gas)

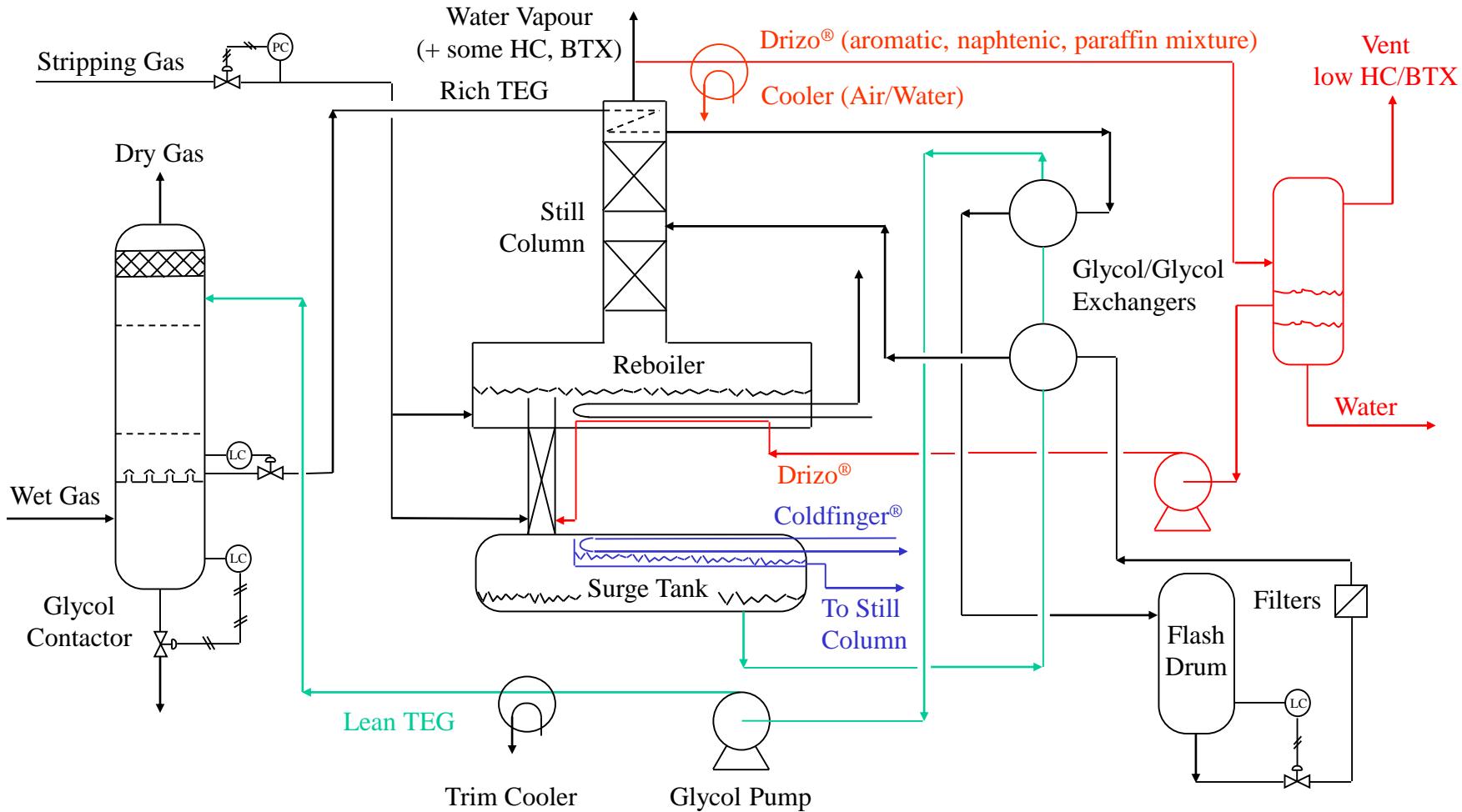


Typical PFD

Gas may be dehydrated to pipeline specifications by glycol absorption. In the liquid state, water molecules are highly associated because of hydrogen bonding. The hydroxyl groups in glycols form similar associations with water molecules. This explains why glycols are such good absorbers of water. Almost all glycol contactors use TEG – triethylene glycol.



TEG Dehydration/Regeneration



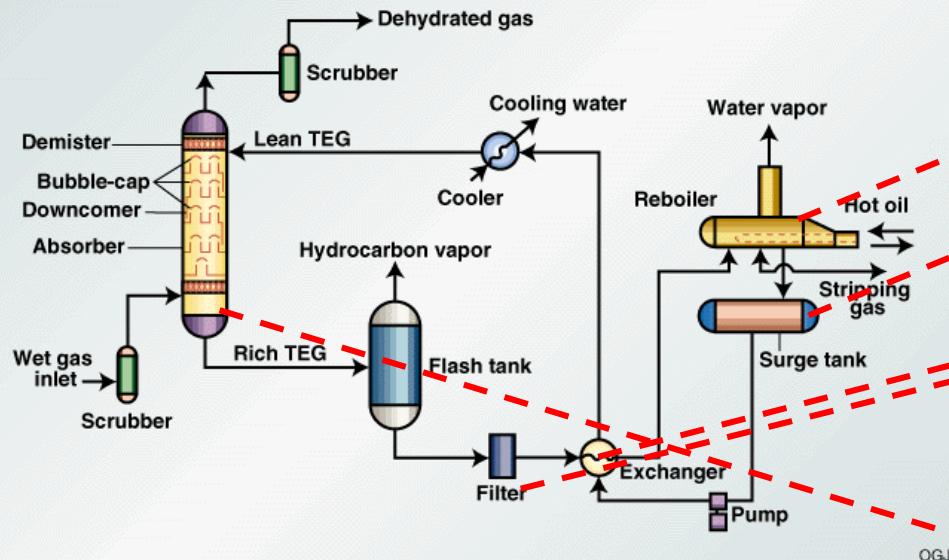
Glycol Dehydration / Regeneration

Glycol Regeneration Methods:

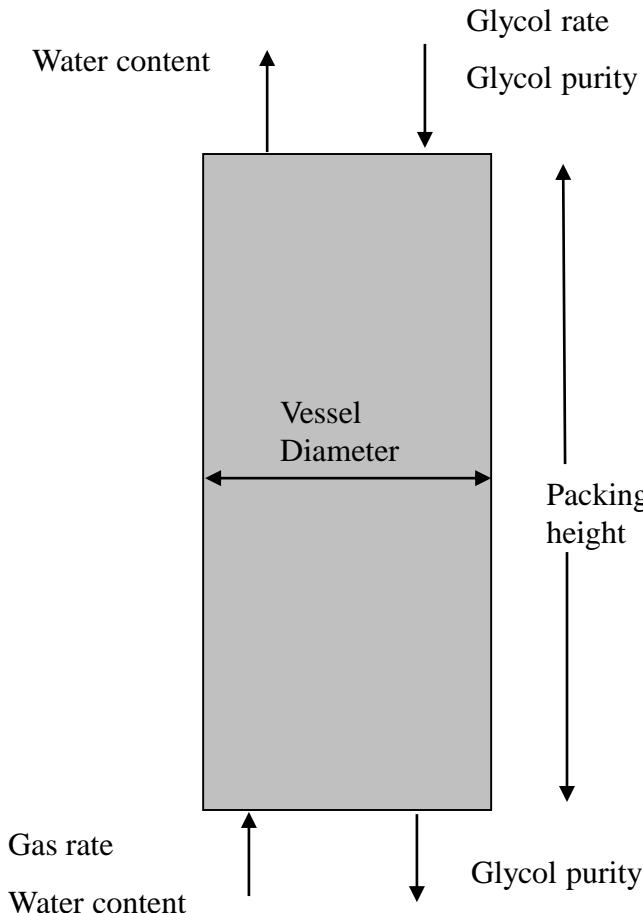
- Reboiler/regenerator column using stripping gas (fuel gas)
- DRIZO®-process using an aromatic recycling stripping solvent
 - little or no venting of hydrocarbons
 - glycol concentrations of > 99.99 wt% can be achieved
 - condensation and recovery of the aromatic hydrocarbons from the still column overhead
- Coldfinger® using a cooling element in the surge tank to condense water, thus reducing the water partial pressure in the vapour space and increasing the lean glycol concentration
 - glycol concentrations of > 99.4 wt% can be achieved without the use of a stripping gas

Glycol System

TEG DEHYDRATION UNIT



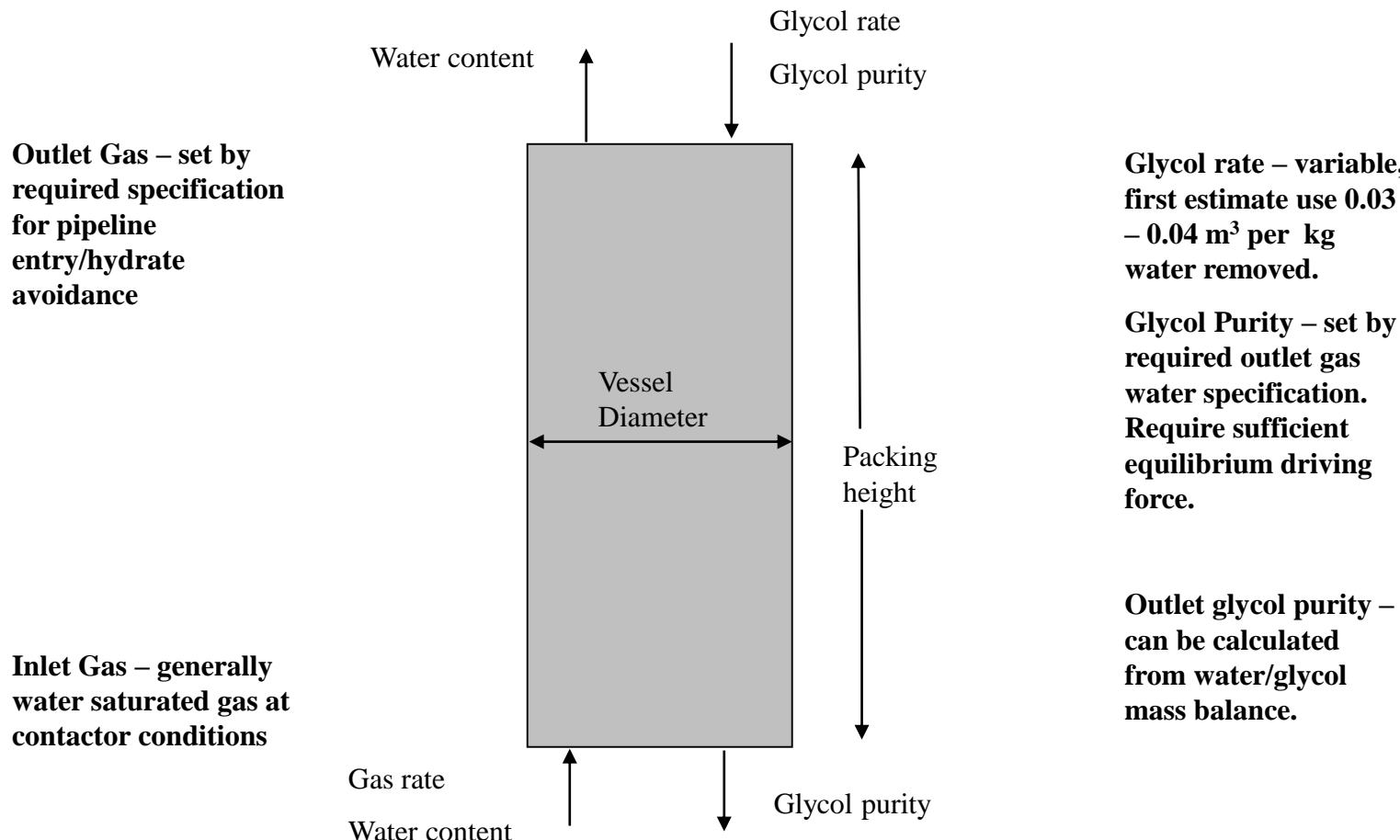
Glycol Contactor Design



Multiple variables

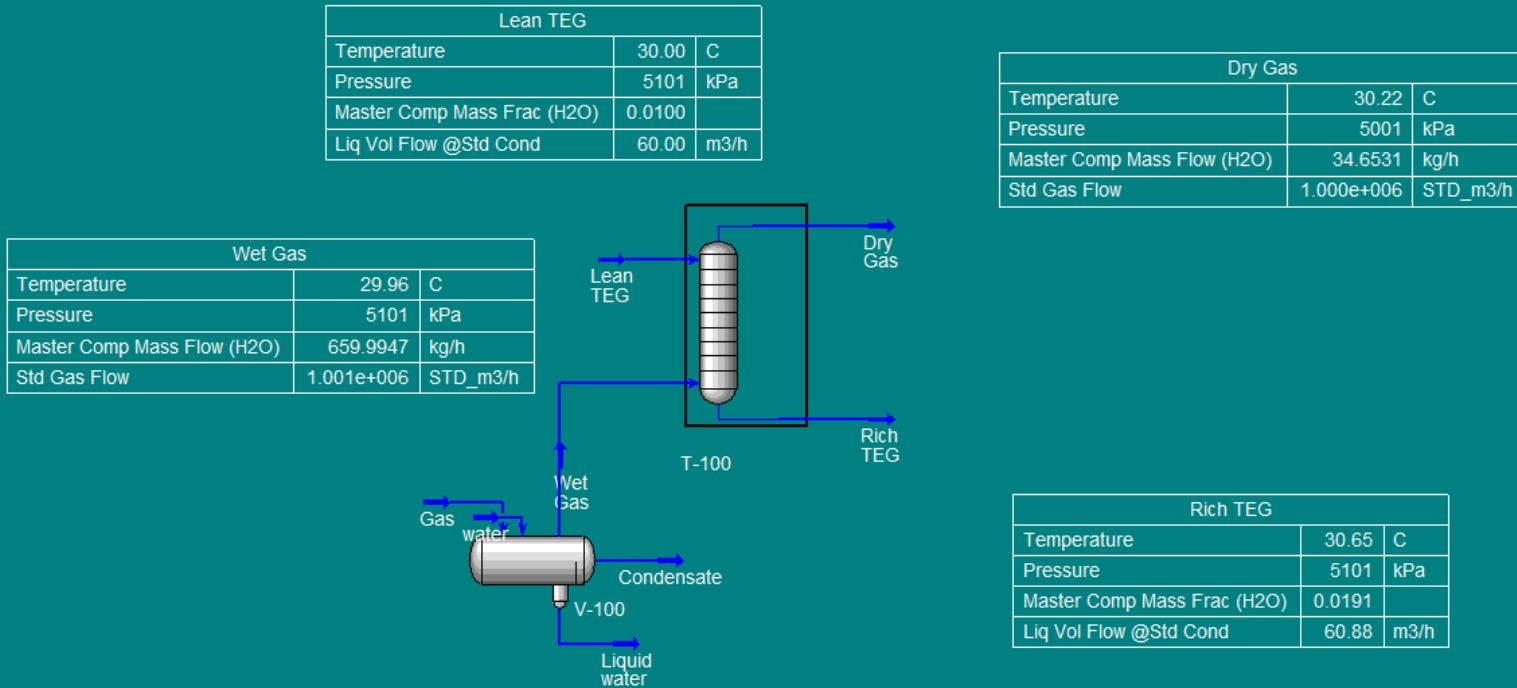
- Know gas rate and required outlet water content
- Outlet water quality sets inlet (lean) glycol purity
- Glycol circulation rate set by equilibrium calculations
- Tower diameter normally established by entrainment limits and flooding
- Packing height from packing characteristics
- Reboiler duty from heat balance - latent and sensible.

Glycol Absorption



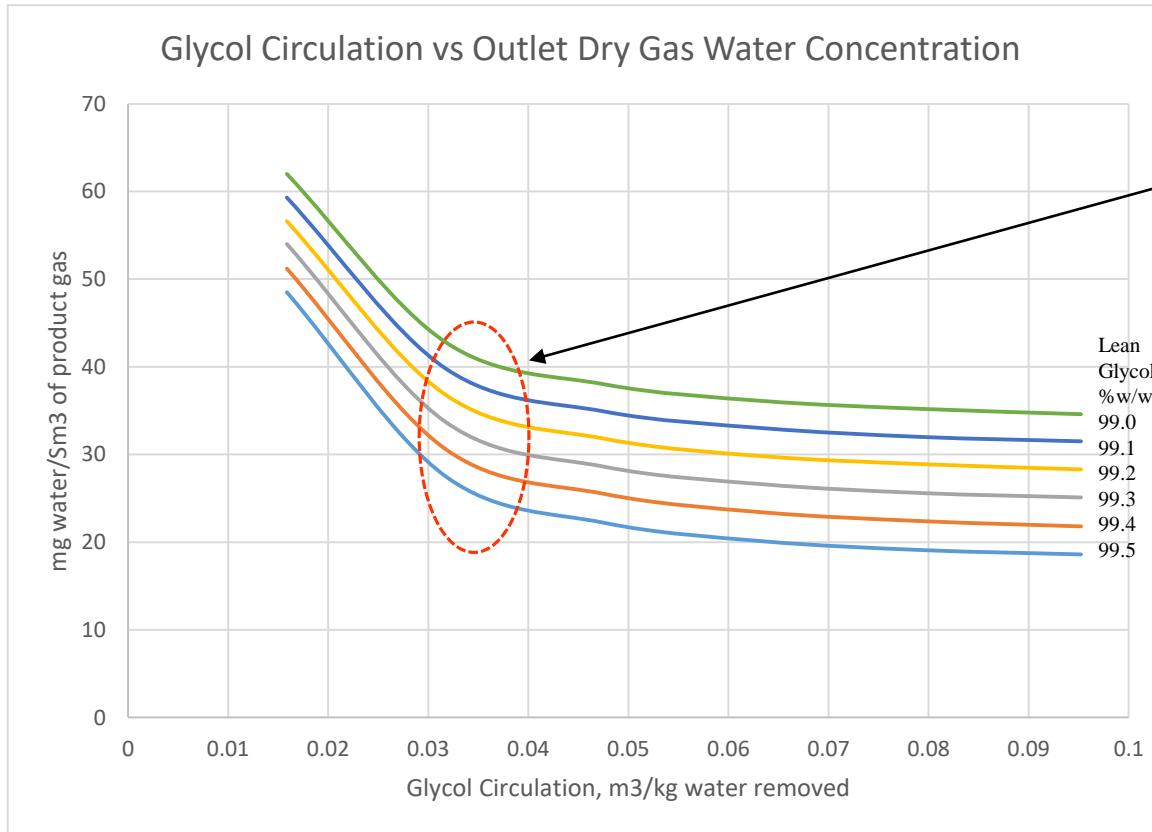
Glycol Circulation Rate

The effect of glycol circulation rate on the Dry Gas water content can be explored using UniSim/Hysys. A glycol absorber is built and sensitivities carried out.



Glycol Circulation Rate

Glycol Circulation sensitivities are shown below.



Benefits of additional glycol circulation marginal.

Water Dewpoint / Lean TEG Concentration

The achievable water dewpoint from the glycol contactor depends on the inlet lean TEG concentration (equilibrium)

It is normal practice to take the desired dewpoint 5-10°C below the equilibrium dewpoint (approach) as equilibrium will not be reached in a contactor.

Problem:

What should be the concentration of a lean TEG solution achieving an equilibrium dewpoint of -15°C in a contactor operating at 30°C ?

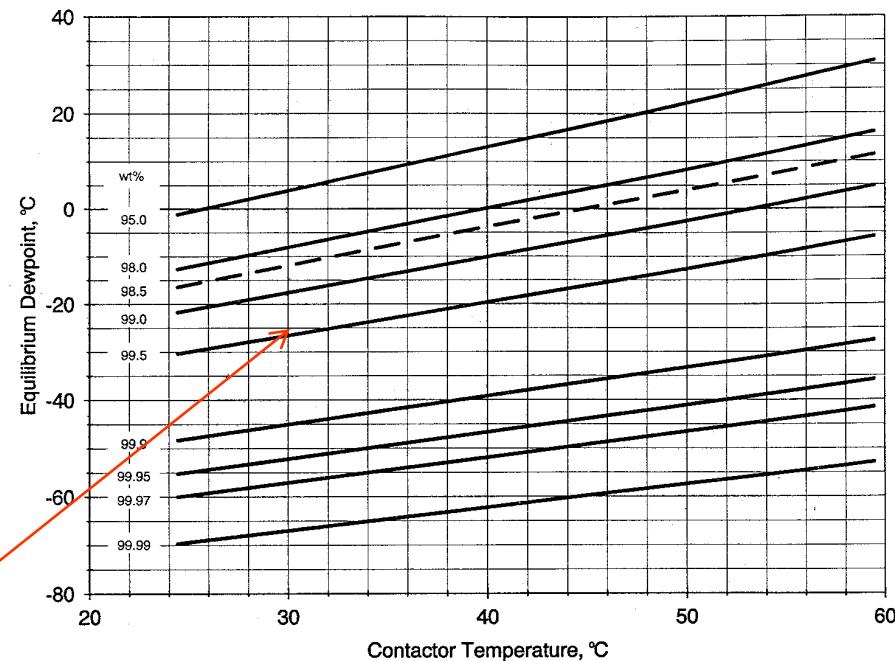
Solution:

As a safety margin, an approach of 10°C is subtracted
→ Equilibrium dewpoint is -25°C

From the chart on the right, read the lean TEG concentration at the intersection of the dewpoint line and the contactor temperature line:

Lean TEG concentration: 99.4 wt%

TEG Equilibrium



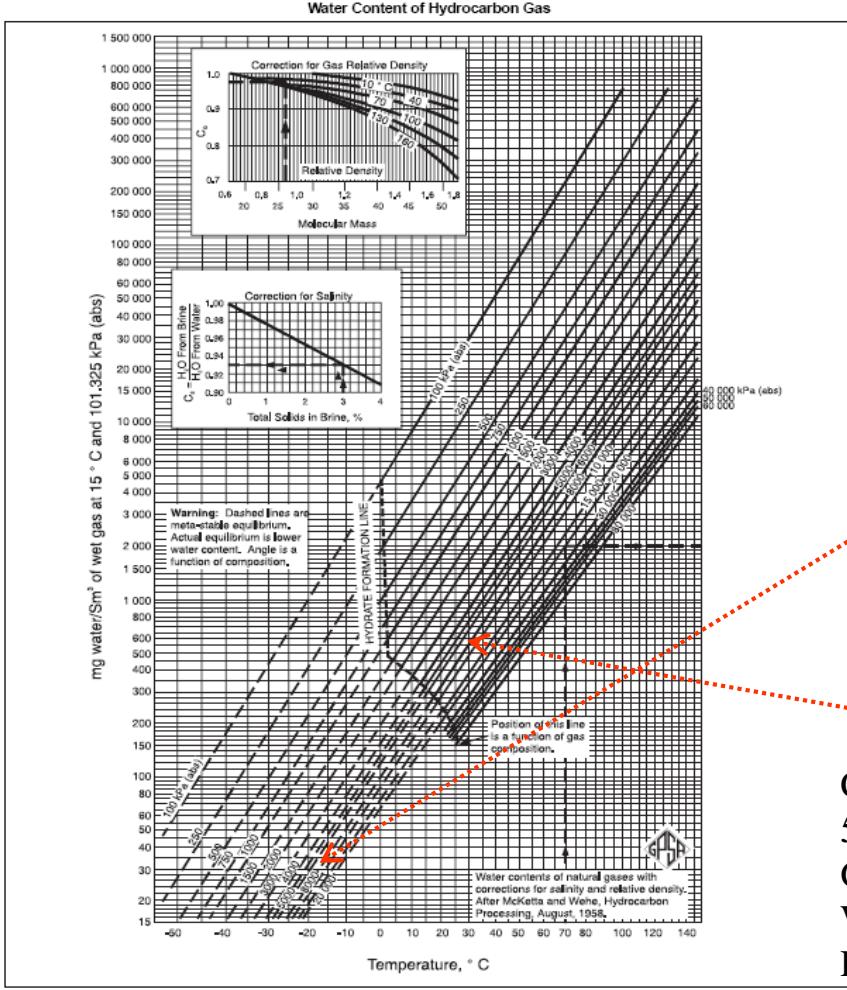
Dewpoint is alternative means for identifying the required water specification.

Water Glycol Equilibrium

The previous chart can be used with the natural gas water saturation chart to predict the equilibrium water vapour content of a gas in contact with TEG.

1. Use the contactor operating temperature to estimate the dewpoint for a known lean TEG concentration.
2. Use the dewpoint temperature at the contactor pressure to estimate the equilibrium water vapour content in mg/m³.

Water Glycol Equilibrium

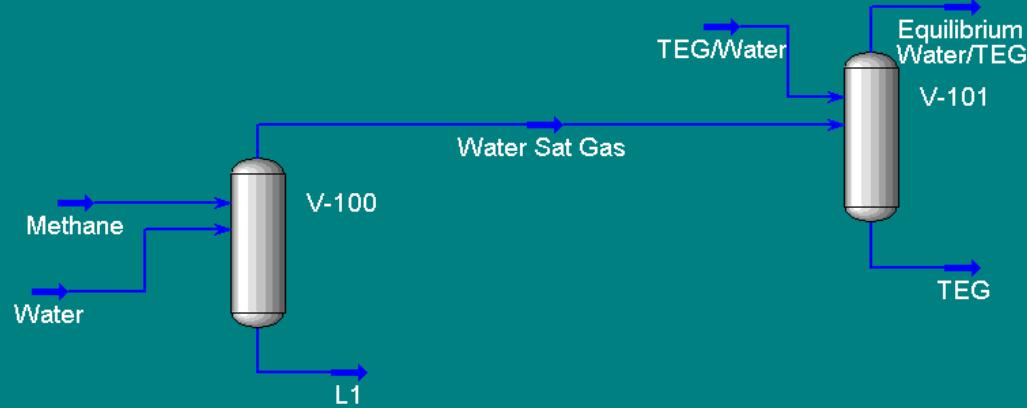


Water Glycol Equilibrium

UNISIM Comparison

TEG/Water		
Temperature	30.00	C
Pressure	80.00	bar
Master Comp Mass Frac (TEGlycol)	0.9900	

Temperature	30.21	C
Pressure	80.00	bar
Std Gas Flow	2360	STD_m3/h
Master Comp Mass Flow (H ₂ O)	0.0850	kg/h

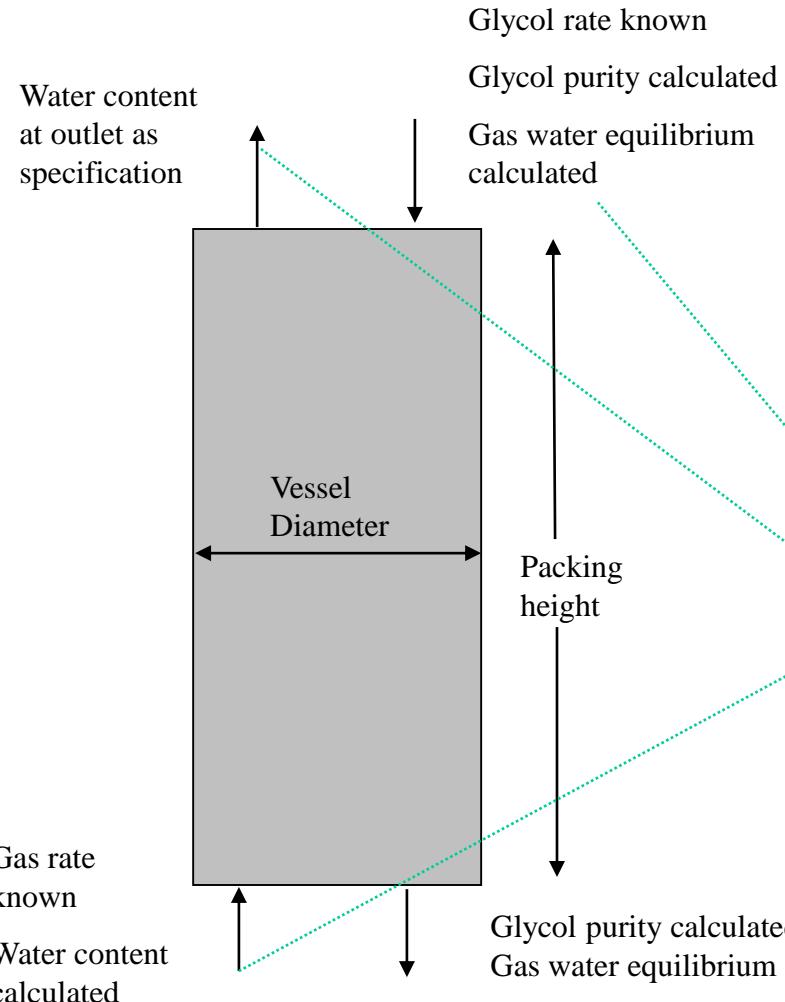


Water Sat Gas		
Temperature	29.96	C
Master Comp Mass Flow (Methane)	1604.2829	kg/h
Std Gas Flow	2366	STD_m3/h
Master Comp Mass Flow (H ₂ O)	1.2824	kg/h
Pressure	80.00	bar

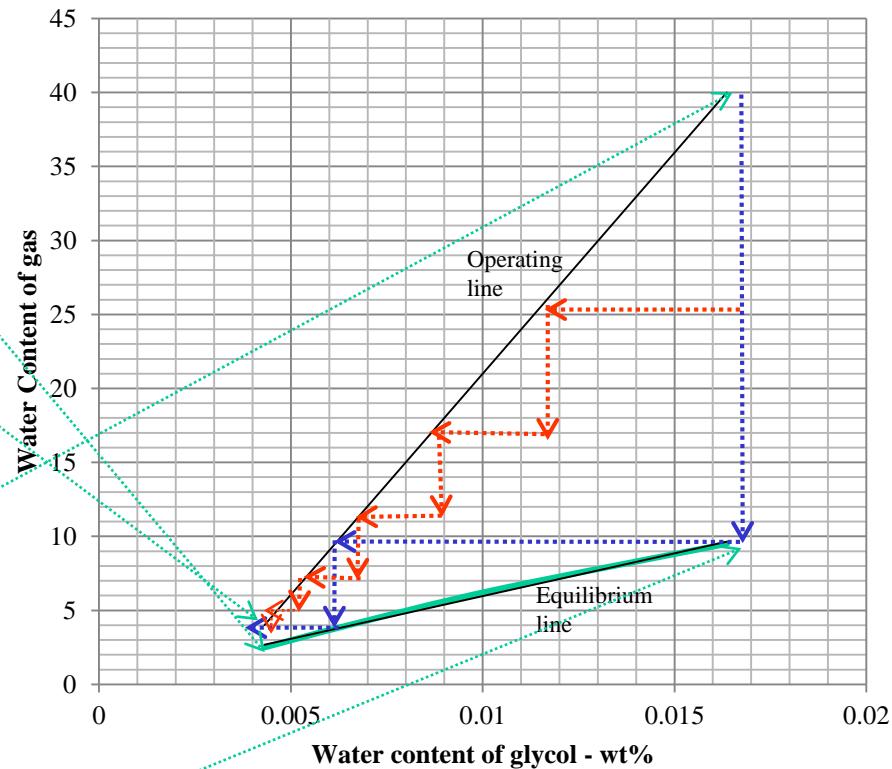
36 mg/Sm3 compare with
35 mg/sm3 from GPSA
chart

540 mg/Sm3 compare with
550 mg/sm3 from GPSA
chart

TEG Absorption



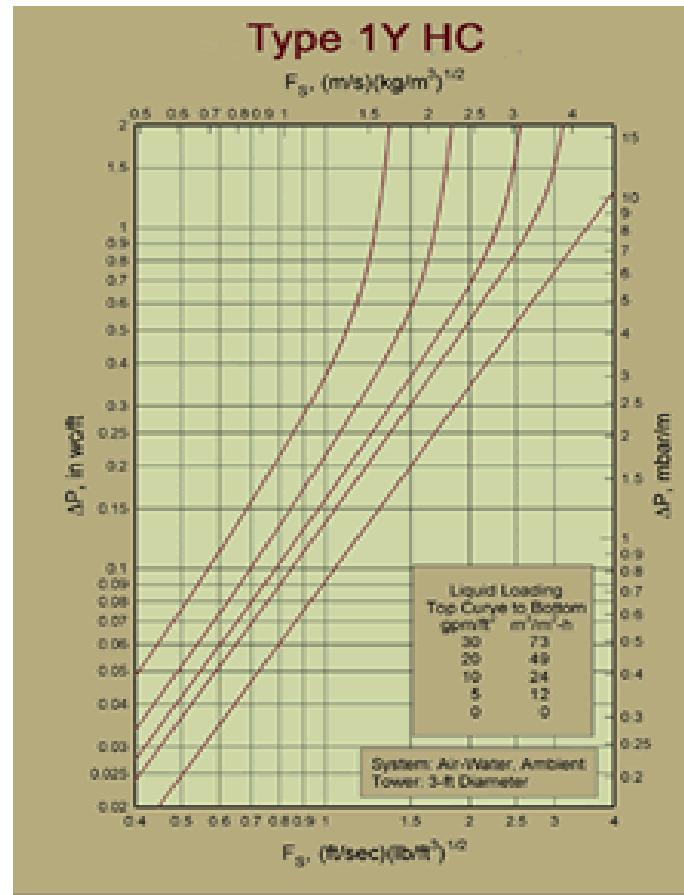
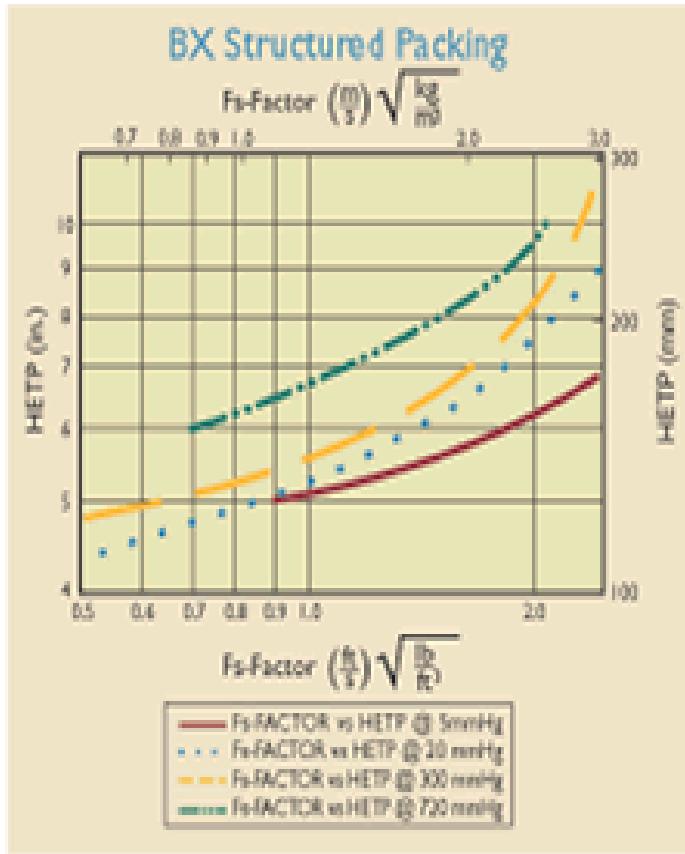
Assuming 50% tray efficiency



Packing Characteristics Supplied by Manufacturer

Structured packing vendors frequently quote an F_s value, $F_s = v \cdot \rho_v^{0.5}$

HETP for a glycol absorber as approximately 1.5 m (GPSA)



Packed Towers

Flood = $f(L, G, \text{packing}, \rho_L, \rho_G, v_G)$

$$X = (L/G) \cdot (\rho_G/\rho_L)^{0.5}$$

$$Y = G^2 \cdot F \cdot v_L^{0.1} / (\rho_G \cdot (\rho_L - \rho_G))$$

L = Liquid mass velocity

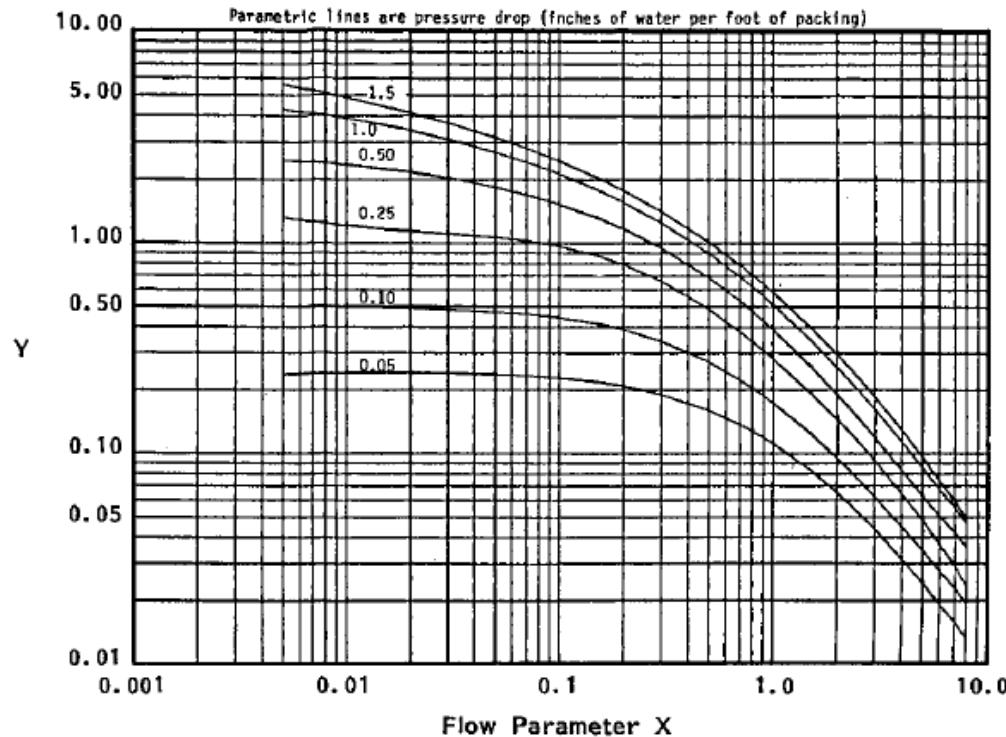
G = Gas mass velocity

F = Packing factor - vendor supplied value depending on packing type and size

ρ_L = Liquid density

ρ_G = Gas density

v_L = Liquid viscosity centistokes



The acceptable pressure drop is given. X and Y are known, G can be calculated. Obtain column diameter from G , Diameter = $(4 \cdot M / \pi \cdot G)^{0.5}$. M is gas mass rate.

Or, if X and Y are known, the pressure drop can be estimated.

Diameter and Flooding

Name	Size or number	Area, m ² /m ³	% voids*	Packing factor, m ⁻¹		
				Normal F _p [†]	Dry F _{pd} [‡]	Vendor
Metals, corrugated sheets						
Mellapak	125Y	125	99	33		Sulzer
	170Y	170	99	39		
	2Y	223	99	46		
	250Y	250	98	66		
	350Y	350	98	75		
	500Y	500	98	112		
	750Y	750	97			
	125X	125	99	16		
	170X	170	99	20		
	2X	223	99	23		
	250X	250	98	26		
	500X	500	98	82		
Mellapak Plus	202Y		99			Sulzer
	252Y	250	98	39		
	352Y		98			
	452Y	350	98	69		
	752Y	500	98	131		
Flexipac	700Y	710	96			Koch-Glitsch
	500Y	495	97			
	1Y	420	98	98	(105)	
	350Y	350	98			
	1.6Y	290	98	59		
	250Y	250	99			
	2Y	220	99	49	(36)	
	2.5Y	150	99			
	3.5Y	80	99	30	(15)	
	4Y	55	99	23	(10.5)	
	1X	420	98	52		
	350X	350	98			
	1.6X	290	98	33		
	250X	250	99			
	2X	220	99	23		
	2.5X	150	99			
	3X	110	99	16		
	3.5X	80	99			
	4X	55	99			
Flexipac High-capacity	700	710	96	223		Koch-Glitsch
	500Z	495	97			
	1Y	420	98	82		
	350Y	350	98			
	1.6Y	290	99	56		
	250Y	250	99			
	2Y	220	99	43		
Intalox	1T	310	98	66		Koch-Glitsch
	1.5T	250	99			
	2T	215	99	56		
	3T	170	99	43		

Tower Diameter

Calculate the diameter of a glycol absorber packed with a mass transfer material which has a packing factor of 30.

The density of the gas and glycol are 80 and 1150 kg/m³ respectively. The gas flowrate is 1.4×10^6 std m³/day. The glycol rate is 2.5 m³/hr, viscosity is 16 centistokes. The gas molecular weight is 18. System pressure is 50 bara, temperature is 30 °C. Z = 0.85.

The acceptable pressure drop for the absorber is 21mm/m packing.

$$\text{Glycol mass rate, } L_m = 2.5 \times 1150/3600 = 0.8 \text{ kg/s}$$

$$\begin{aligned} \text{Gas mass rate, } G_m &= (1.4 \times 10^6 \times 18)/(23.96 \times 24 \times 3600) \quad (1\text{kgmol} = 23.96 \text{ s m}^3) \\ &= 12.2 \text{ kg/s} \end{aligned}$$

$$X = (L/G) \cdot (\rho_G / \rho_L)^{0.5} = (0.8/12.2) \times (80/1150)^{0.5} = 0.017$$

Tower Diameter

$$Y = C \cdot G^2 \cdot F \cdot v_L^{0.1} / (\rho_G \cdot (\rho_L - \rho_G))$$

$$= 0.85$$

$$G^2 = 0.85 \times 80 \times (1150 - 80) /$$

$$(10.764 \times 30 \times 16^{0.1})$$

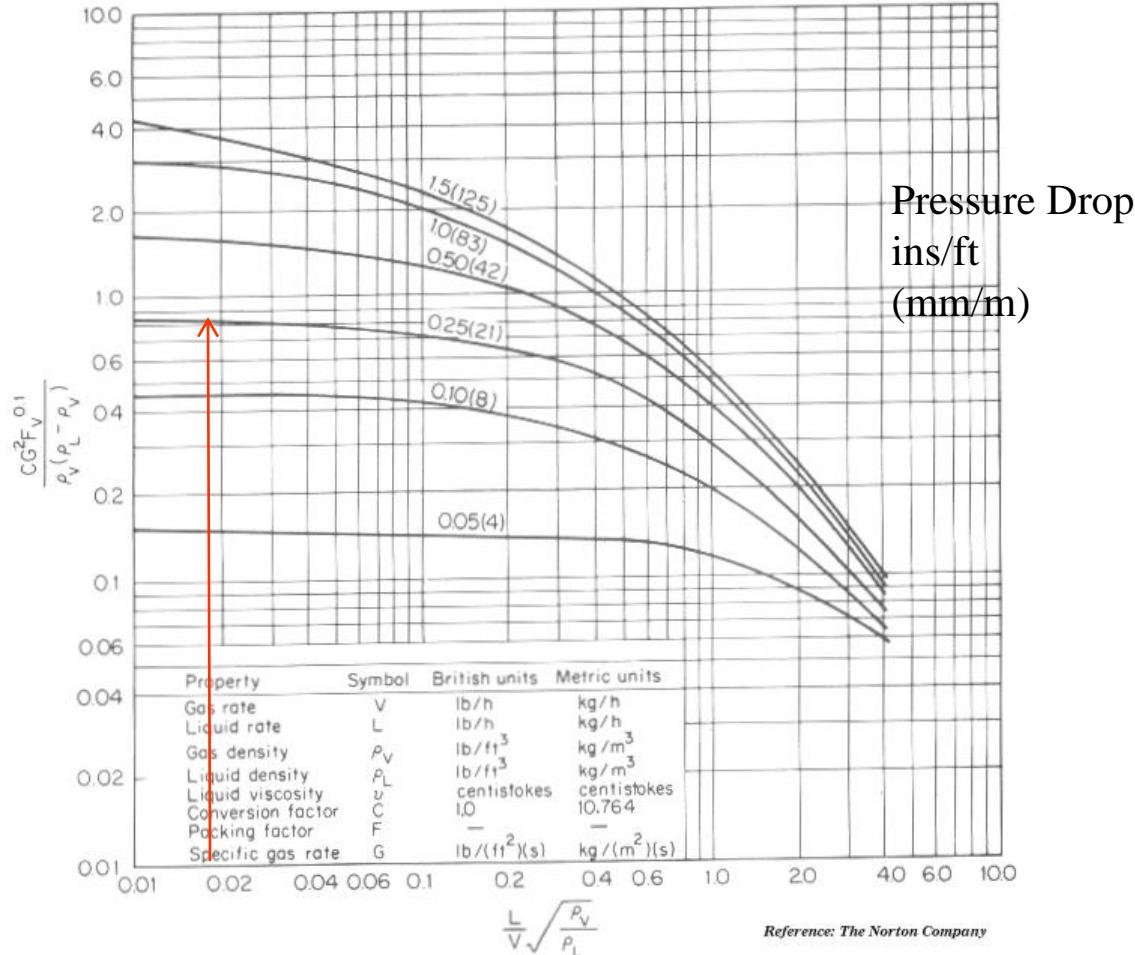
$$= 170$$

$$G = 13 \text{ kg/m}^2\text{.s}$$

$$G = Gm/\text{Tower CSA}$$

$$= Gm/(3.14 \times D^2/4)$$

Solving for D = 1.16m



Tower Diameter

Check for maximum entrainment velocity;

$$v_{\max} = K_s \cdot \left(\frac{r_L - r_G}{r_G} \right)^{0.5}$$

v_{\max} Maximum allowable superficial gas velocity (m/s)

K_s Constant (m/s)
0.0488 for 24" tray spacing and 2" seal

over bubble cap slots

0.1 for structured packing

ρ_L Density of TEG (kg/m³)

ρ_G Density of gas (kg/m³)

Calculate gas volume flow;

$$\begin{aligned} \text{Volume flow, } V_g &= 1.4 \times 10^6 \text{ std m}^3/\text{day} \\ \text{Actual volume flow} &= 1.4 \times 10^6 \times Z \times 1/P \times T/288 \\ &= 0.025 \times 10^6 \text{ m}^3/\text{day} \\ &= 0.29 \text{ m}^3/\text{s} \end{aligned}$$

$$\text{Gas velocity} = 0.29/\text{CSA} = 0.29/1.06 = 0.27 \text{ m/s}$$

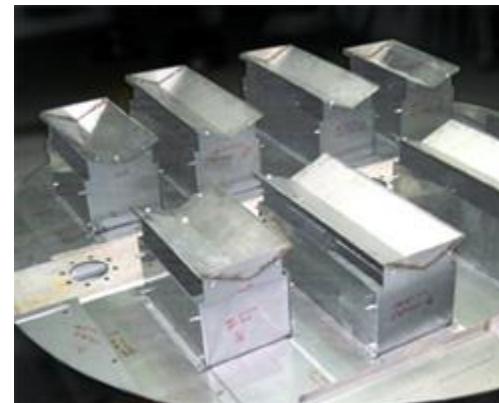
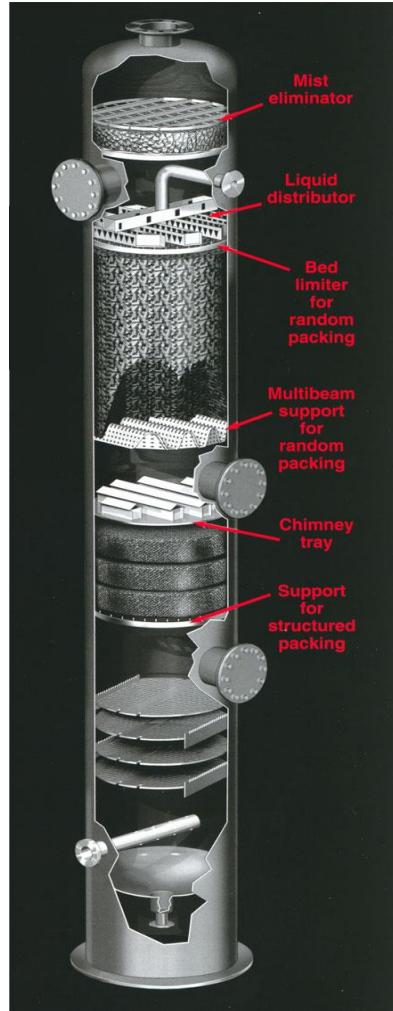
Compare with v_{\max}

$$V_{\max} = 0.1 \times ((1150 - 80)/80)^{0.5} = 0.36 \text{ m/s}$$

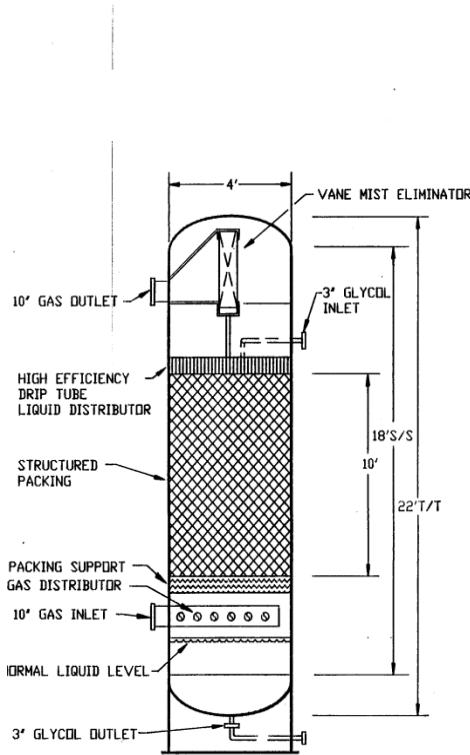
Hence, tower acceptable from entrainment criteria.

Distributors / Collectors

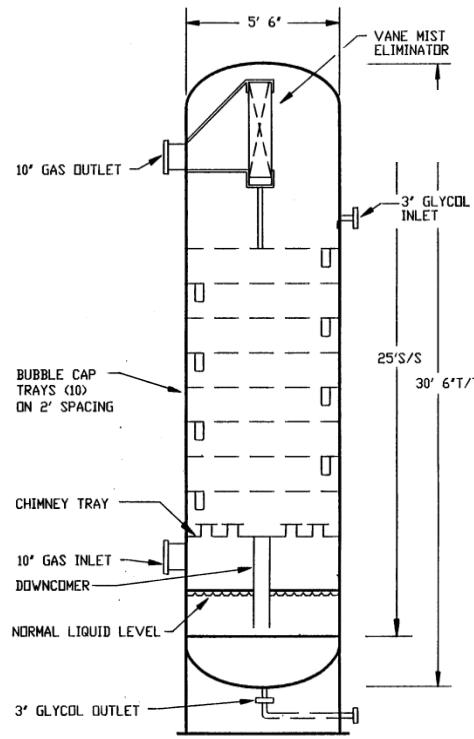
Column internal components are required to uniformly distribute the fluids and efficiently collect liquid from the packing/trays.



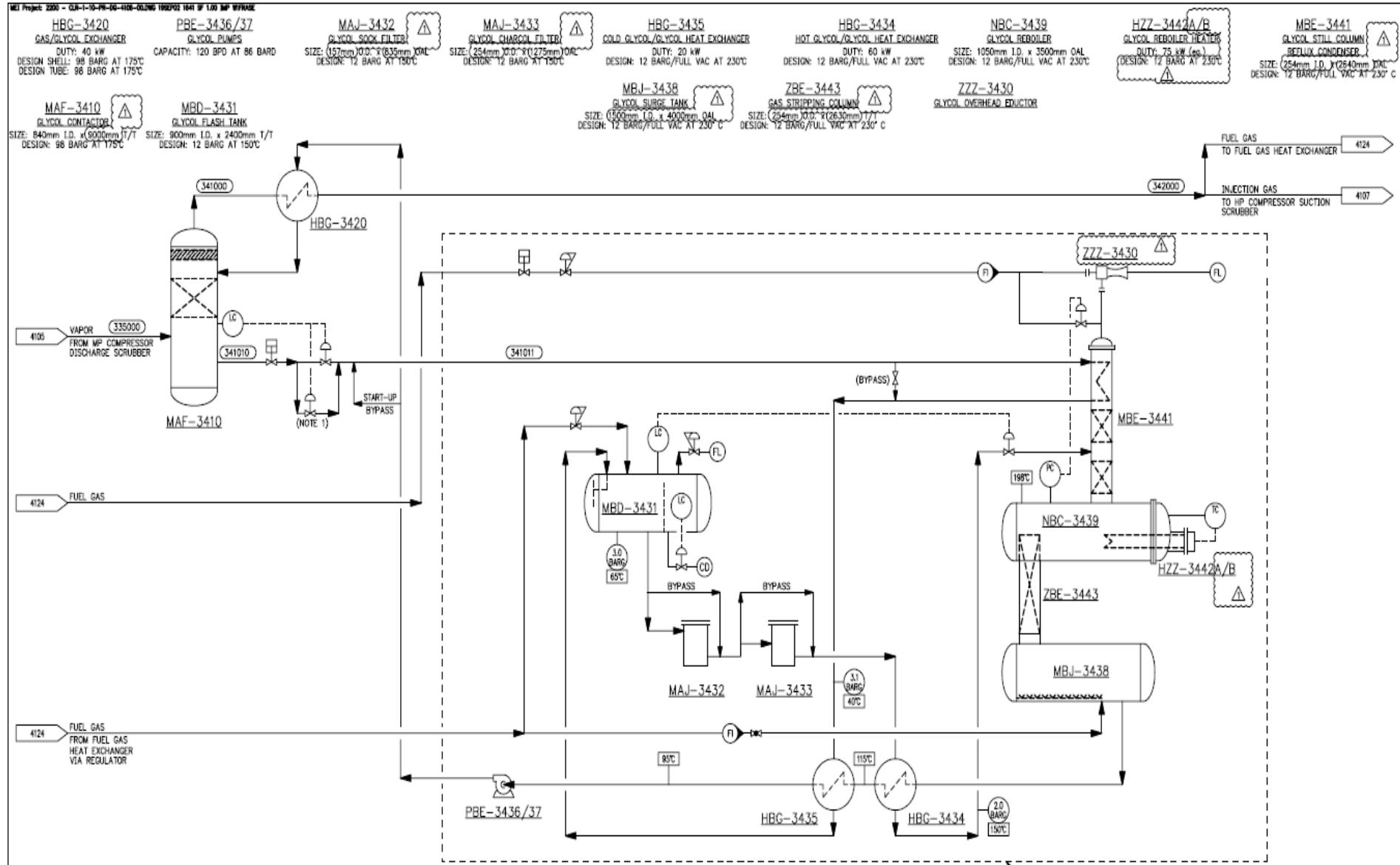
Tray – Packing Comparison



238 GLYCOL CONTACTOR
WITH STRUCTURED PACKING

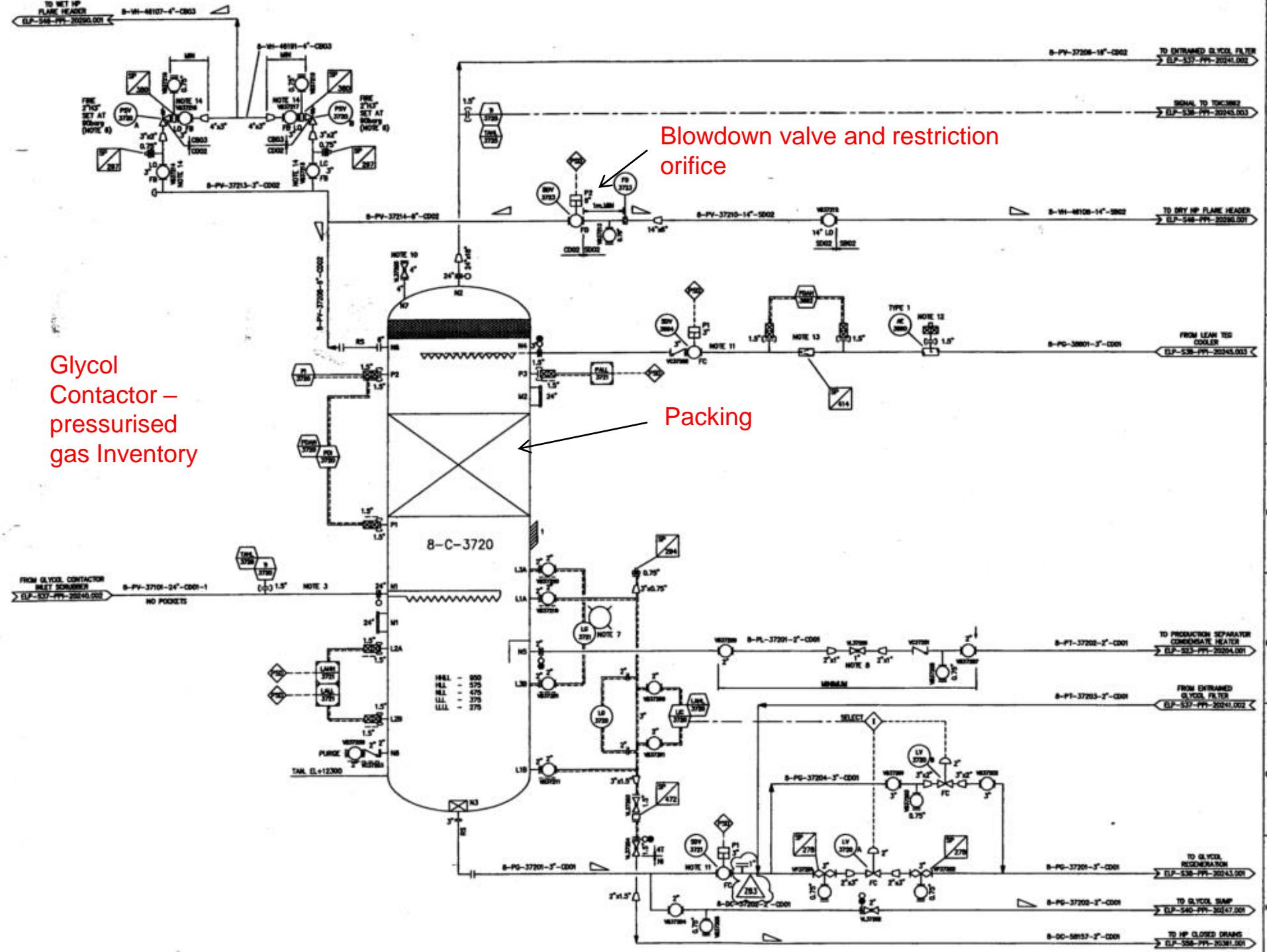


TRAYED GLYCOL CONTACTOR



Year	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Year 14	Year 15	Year 16	Year 17	Year 18	Year 19	Year 20
Stream	335000	340100	340101	340102	340200	335000	340100	340101	340200	335000	340100	340101	340200	335000	340100	340101	340200	335000	340100	
Vapor Fraction	1	1	0	0.02	35.4	36.8	30	34	35.4	36.8	30	34	35.4	36.8	30	34	35.4	36.8	30	34
Temperature (°C)	35	35.4	35.8	43.2	36.8	30	34	35.4	36.8	30	34	35.4	36.8	30	34	35.4	36.8	30	34	
Pressure (bar)	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	65.5	
Density (kg/m³)	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	
Viscosity (cP)	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	0.003	
Dew (100% Aqueous)	11.83	11.83	104	11.83	37.42	37.36	1077	37.38	49.24	48.2	100	49.2	41.83	41.76	100	49.2	41.83	41.76	100	
Mass Flow (kg/h)	132001	131844	7698	131844	411225	411201	7897	41090	53436	53396	7966	53396	40955	40952	7966	53396	40955	40952	7966	
Mass Heat Capacity (kJ/kg·K)	3.2	3.19	2.2	2.24	3.11	3.15	3.14	2.29	2.33	3.12	3.11	2.33	2.37	3.1	3.21	3.2	3.2	2.35	3.16	
Molecular Weight	22.36	22.38	130	22.38	22.03	112.62	112.62	22.03	21.75	21.75	107.13	107.13	21.75	22.38	22.38	106.64	106.64	22.39		
Compressibility	0.7	0.7	0.72	0.71	0.72	0.72	0.72	0.72	0.72	0.72	0.73	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	
k (C/P)	1.83	1.83	1.05	1	1.78	1.81	1.8	1.85	1	1.79	1.78	1.78	1.05	1	1.77	1.84	1.83	1.05	1	
Heat of Vaporization (kJ/kg)	95	91	305	81	65	62	401	82	81	81	403	63	76	76	402	79	402	79		
Standard Density (kg/m³)			1111		1112			1113			1112									

ZZZ-3400
GLYCOL REGENERATION PACKAGE



Upthrust Damage



Glycol System Operating Issues

Hydrocarbon Content - If hydrocarbons are present, the unit may be subject to foaming and the formation of gummy or tarry deposits in the reboiler, heat exchangers, and absorber.

Corrosion - A glycol pH of less than 6 indicates a potentially serious corrosion problem with the system.

Foaming - Can be induced by hydrocarbons, salts and degradation products.

Salt precipitation - Can cause problems with reboiler.

Glycol Carryover - Poor tower design

Chimney tray Design - Poor design can lead to re-entrainment.

A simple visual inspection of the glycol can provide clues for identifying many glycol problems. The following conditions can indicate major problems:

- The presence of a finely divided black precipitate is the result of iron corrosion.
- A black viscous glycol solution may be the result of heavy hydrocarbon contamination or glycol polymerization due to thermal breakdown or interaction with hydrocarbons.
- A sweet, aromatic odour may be an indicator of thermal degradation of the glycol.
- A two-phase solution indicates hydrocarbon contamination.

Adsorption

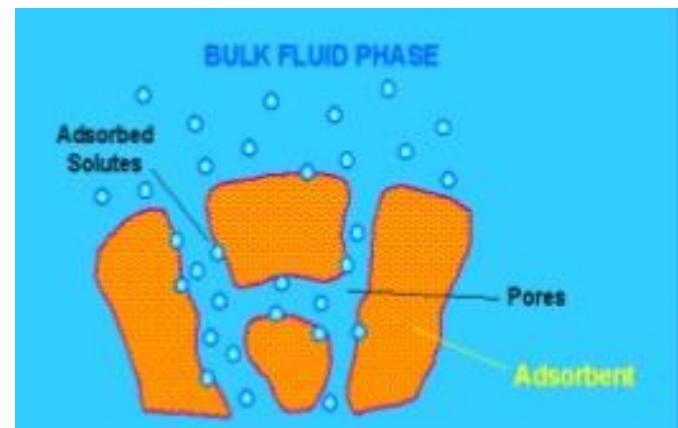
Dehydration with TEG is termed absorption. Adsorption is a phenomenon where a solid surface holds a molecule either by chemisorption or physisorption. The most common industrial application is physisorption where van der Waals forces attract and hold molecules onto a solid surface.

Desiccants in common commercial use fall into one of three categories:

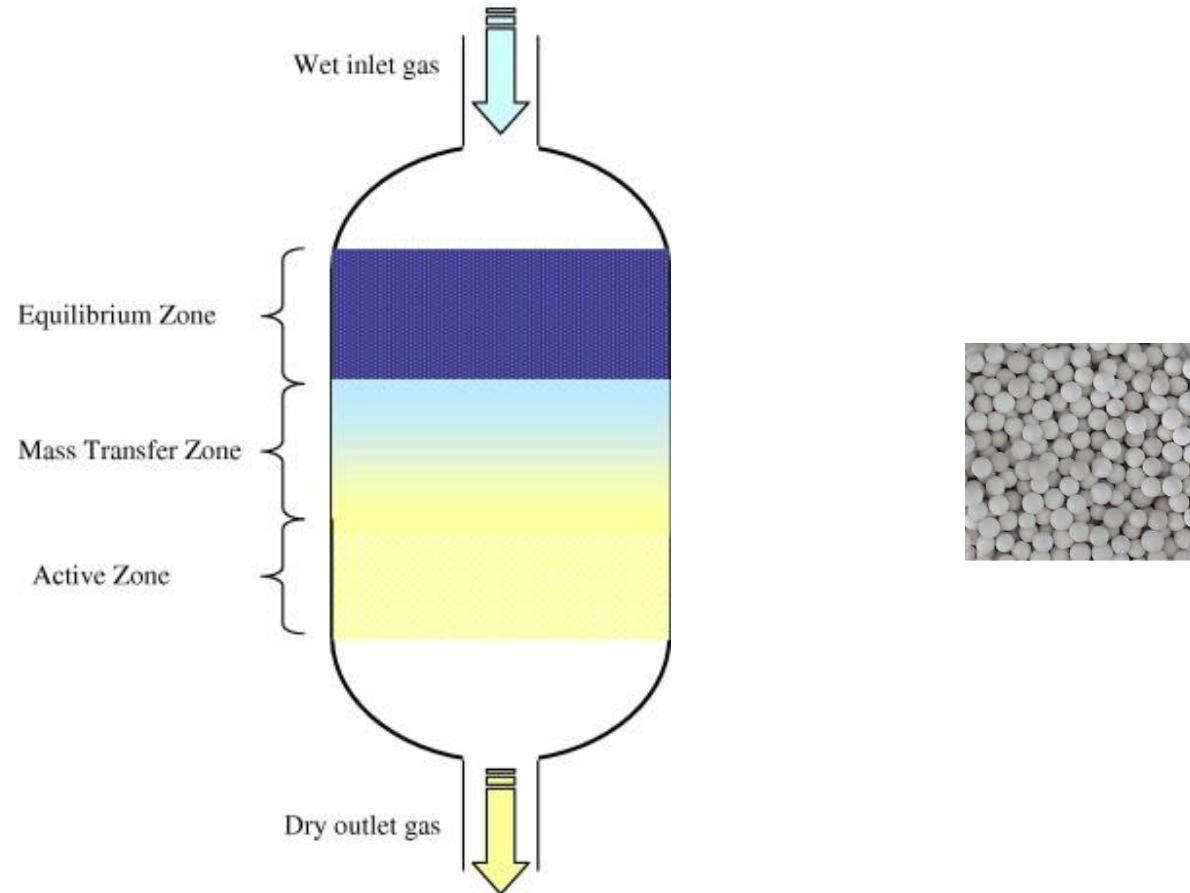
Gels – alumina or silica gels manufactured and conditioned to have an affinity for water.

Alumina – a manufactured or natural occurring form of aluminium oxide that is activated by heating.

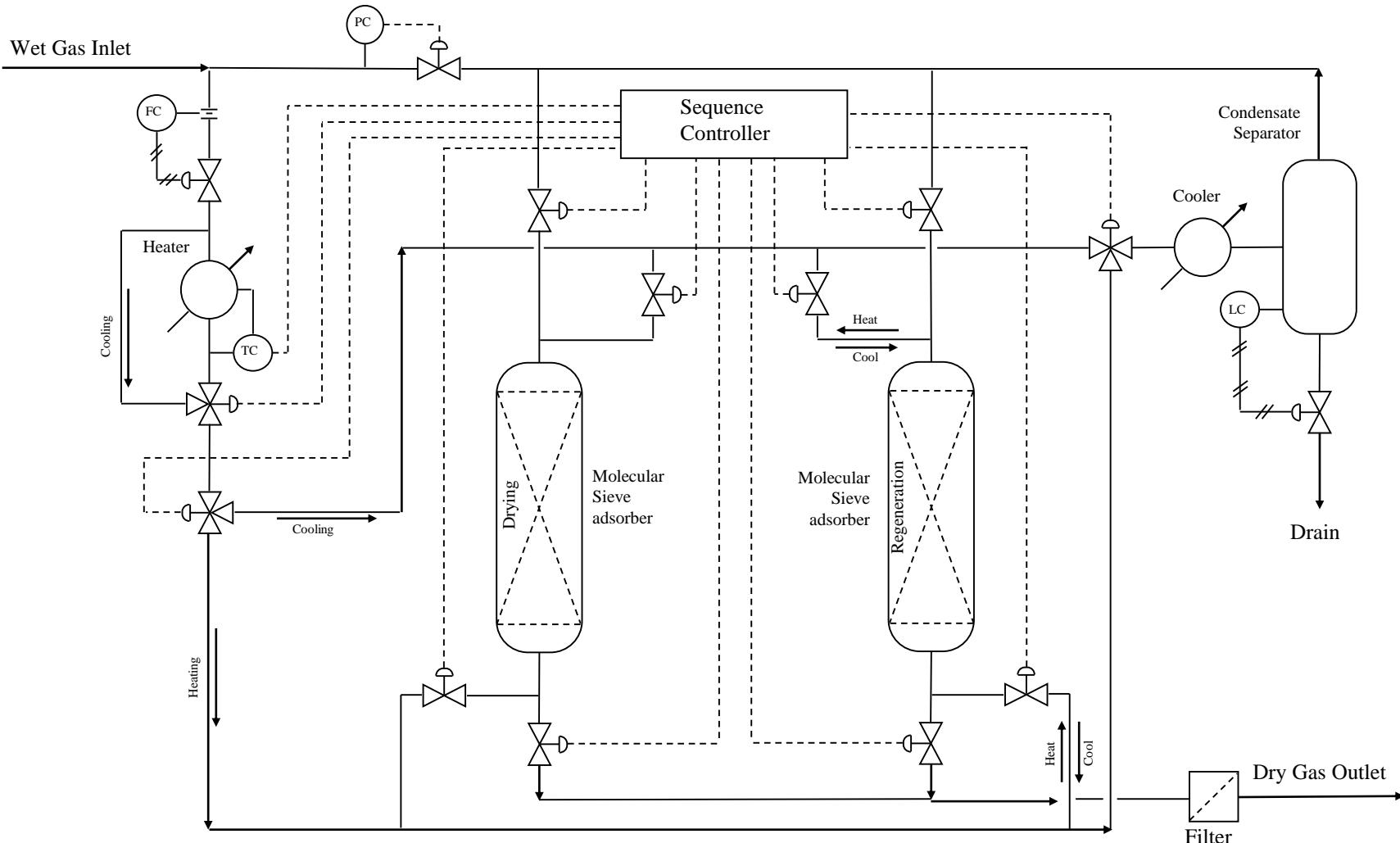
Molecular Sieves – manufactured or naturally occurring aluminosilicates exhibiting a degree of selectivity based on crystalline structure in their adsorption of natural gas constituents.



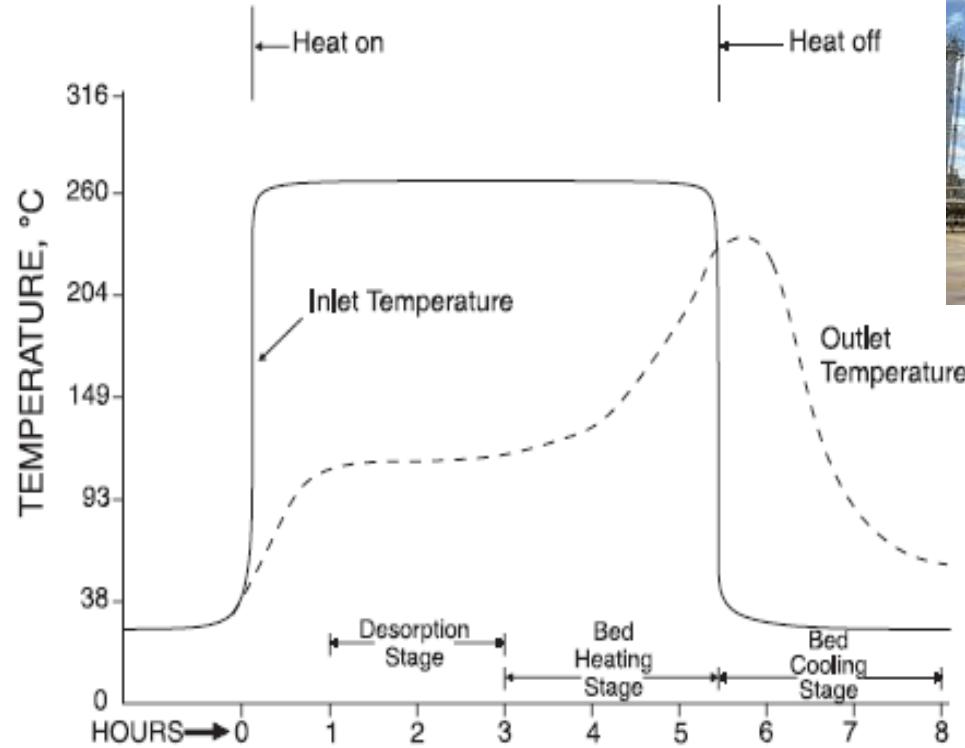
Molecular Sieve System



Molecular Sieve System



Typical Temperature profiles



Molecular Sieve – Preliminary Design

The allowable superficial velocity can be calculated from the accompanying figure.

Once the superficial velocity is estimated the bed cross sectional area and diameter (D) can be readily calculated from the gas volume flowrate.

An absorption cycle time has to be estimated typical 8 – 12 hours.

During the adsorption cycle, the bed is operating in the three previously mentioned zones; The top equilibrium zone where the desiccant in this zone is in equilibrium with the wet inlet gas. The mass transferzone (MTZ) is where the water content of the gas is reduced from saturation to typically < 1 ppmv. Followed by the bottom unused desiccant active zone.

Molecular Sieve – Preliminary Design

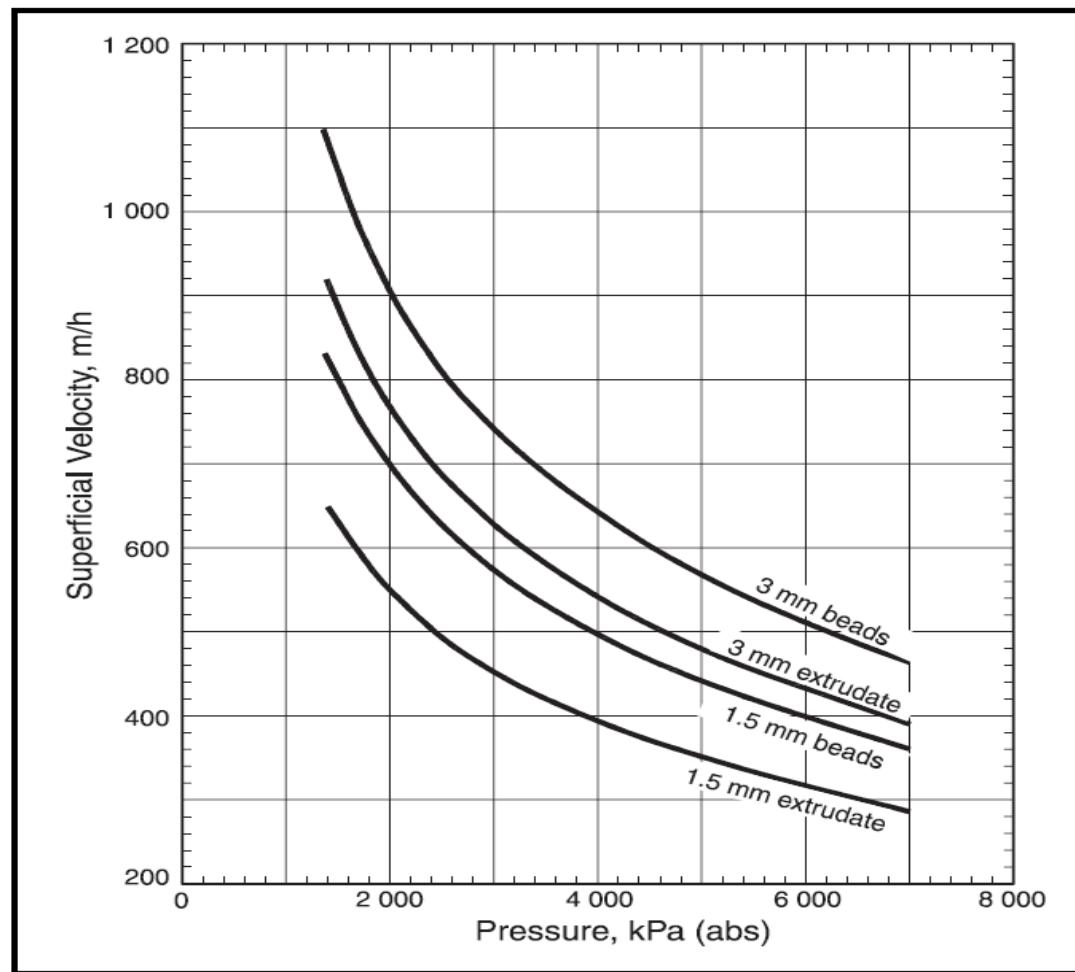
If the bed operates too long in adsorption, the mass transfer zone begins to move

out the bottom of the bed causing a “breakthrough.” At breakthrough, the water content of the outlet gas begins to increase and will eventually reach feed gas water content when the MTZ is completely displaced.

Both water capacity and the rate at which solid desiccants adsorb water decline as the material ages. The object of the design is to install enough desiccant such that after three to five years, the mass transfer zone will be at the bottom of the bed at the end of the adsorption cycle.

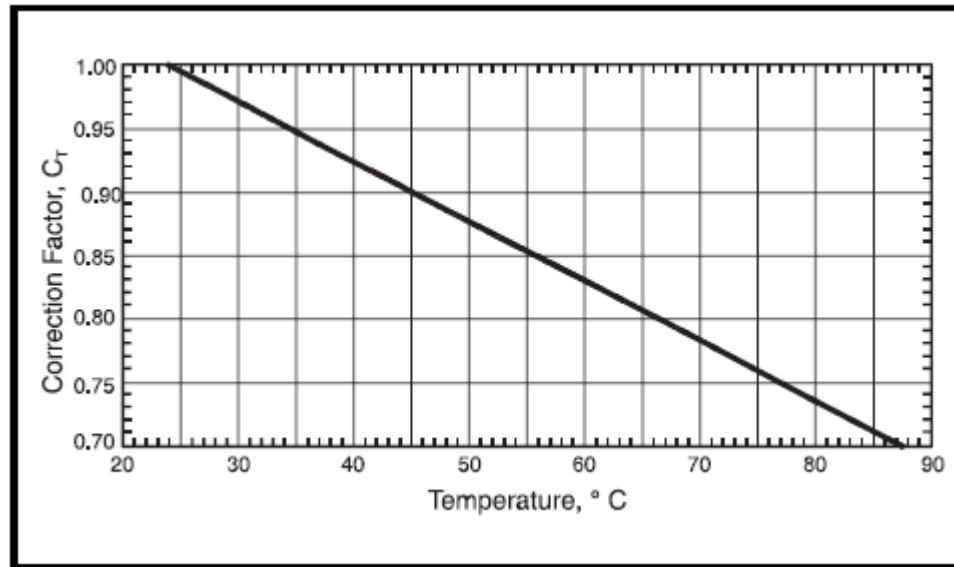
Molecular Sieve – Preliminary Design

Allowable Velocity for Mole Sieve Dehydrator



Molecular Sieve – Preliminary Design

In the saturation zone, molecular sieve is expected to hold approximately 13 kg of water per 100 kg of sieve. New sieve will have an equilibrium capacity near 20kg; 13kg represents the approximate capacity of a 3-5 year old sieve.



Molecular Sieve – Preliminary Design

To determine the mass of desiccant required in the saturation zone, calculate the amount of water to be removed during the cycle and divide by the effective capacity.

$$S_s = W_r / (0.13 \times C_t)$$

S_s = molecular sieve material required in sat zone, kg

W_r = Water removed per cycle, kg

C_t = Temperature correction factor

Calculate the L_s , the length of packed bed in the saturation zone from;

$$L_s = S_s \times 4 / (\pi \times D^2 \times \rho)$$

Sieve density is around 700 kg/m³

Molecular Sieve – Preliminary Design

The length of the mas transfer zone can be calculated from:

$$L_{MTZ} = (V/560)^{0.3} \times K$$

V = Superficial vapour velocity

K = 1.7 for 3mm sieve, 0.85 1.5 mm sieve

The total bed height is the sum of the saturation and MTZ lengths.

A further 1.9m is added top and bottom to provide even gas distribution.

Operating Characteristics

Molecular Sieve Dehydration

- Recommended Operating Range:
 - $T < 50^{\circ}\text{C}$
 - $T >$ Hydrate point
- Desiccant Service:
 - 3 - 5 years in absence of poisoning. Limited by loss of capacity, dusting and breakage
- Dehydration Obtained:
 - Essentially bone dry (water free) gas. Dewpoint: -75 to -125°C (Silica Gel: -60°C)
- Length of Cycle:
 - Varies with water loading and gas rate: 4 - 24 hours
- Regeneration:
 - Temperature: $175 - 300^{\circ}\text{C}$. About 5 to 15% of the total gas stream is used for regeneration. 8 hour cycle: 6 hours heating, 2 hours cooling



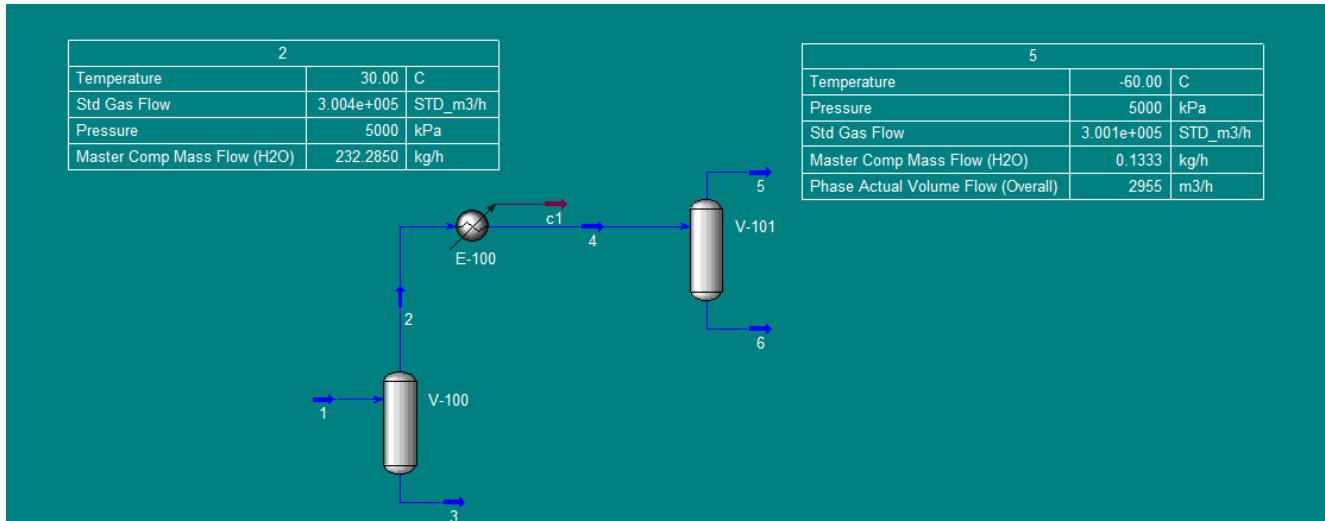
Operating Characteristics

Molecular Sieve Dehydration

- Advantages:
 - Low exit dewpoint
 - Effective dewpoint depression over a wide range of operating conditions
 - Compact
 - Relatively low initial investment for small amounts of gas (batch or semi-batch operation)
 - Rated capacity may be increased by bypassing some wet gas
- Disadvantages:
 - High initial investment
 - Desiccant sensitive to poisoning
 - Rated capacity declines with pressure
 - Pressure drop is higher than liquid desiccant systems
 - Regeneration heat load can be high in relation to the amount of gas processed

Molecular Sieve – Preliminary Design

Water saturated gas is to be dried at 30DegC and 50 bar to a residual water dewpoint of -60 DegC. The gas flowrate is 300000 sm³/hr. Size a molecular sieve vessel for this duty.



From simulation amount of water removed = $232.3 - 0.1 = 232.2 \text{ kg/hr}$

Using a two bed system on an 8 hour cycle time water removed per cycle = $8 \times 232.2 / 3 = 619 \text{ kg}$

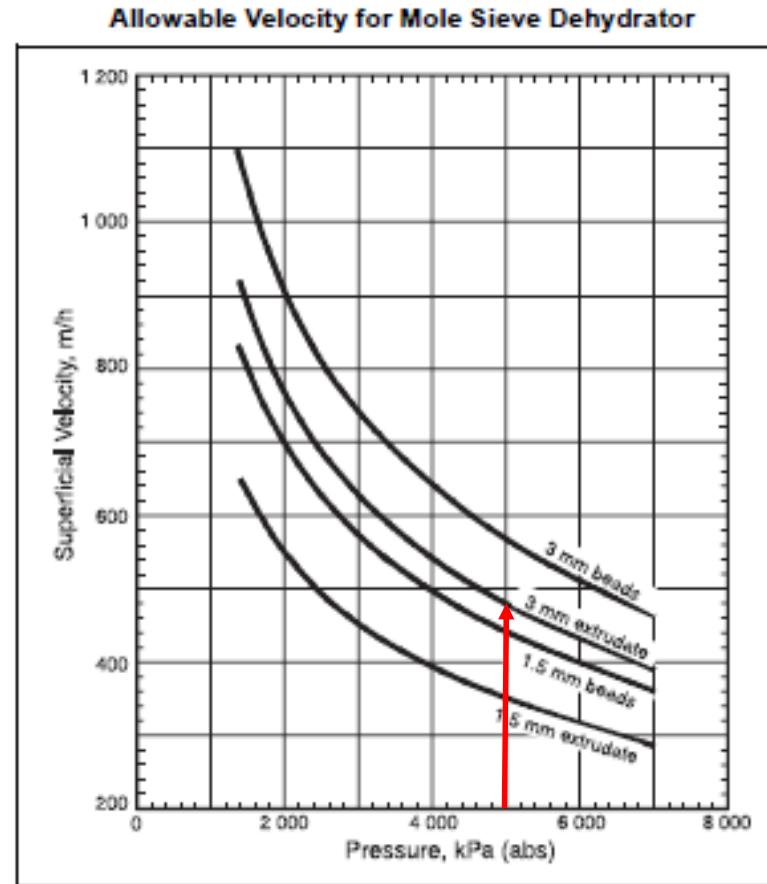
Molecular Sieve Preliminary Design

At 50 bar the superficial velocity is 480 m/hr for 3mm extrusions.

Actual gas volume rate =
 $2955 \text{ m}^3/\text{hr}$

Bed CSA = $2955/480 =$
 6.16 m^2

Bed Diameter = 2.8 m



Molecular Sieve Preliminary Design

$$S_s = W_r / (0.13 \times C_t) = 619 / (0.13 \times 0.97) = 5384 \text{ kg}$$

S_s = molecular sieve material required in sat zone, kg

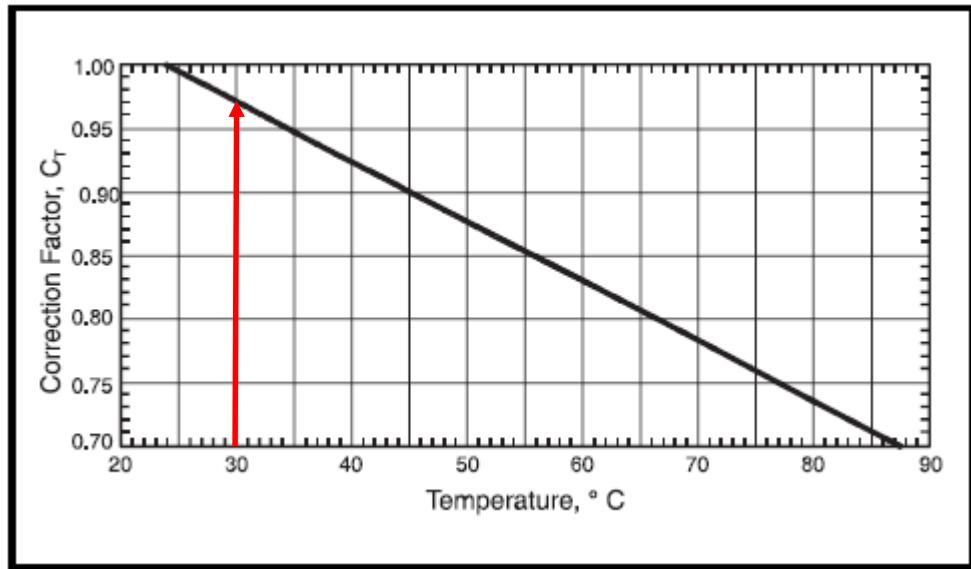
W_r = Water removed per cycle, kg

C_t = Temperature correction factor

Calculate the L_s , the length of saturation zone;

$$L_s = S_s \times 4 / (\pi \times D^2 \times \rho) = 5384 \times 4 / (\pi \times 2.8^2 \times 700) = 1.3 \text{ m}$$

Sieve density is around 700 kg/m³



The length of the mas transfer zone can be calculated from:

$$L_{MTZ} = (V/560)^{0.3} \times K = (480/560)^{0.3} \times 1.7 = 1.62\text{m}$$

V = Superficial vapour velocity

K = 1.7 for 3mm sieve, 0.85 1.5 mm sieve

The total bed height is the sum of the saturation and MTZ lengths.

A further 1.9m is added top and bottom to provide even gas distribution.

$$\text{Vessel Height} = 1.3 + 1.62 + 2 \times 1.9 = 6.7 \text{ m}$$

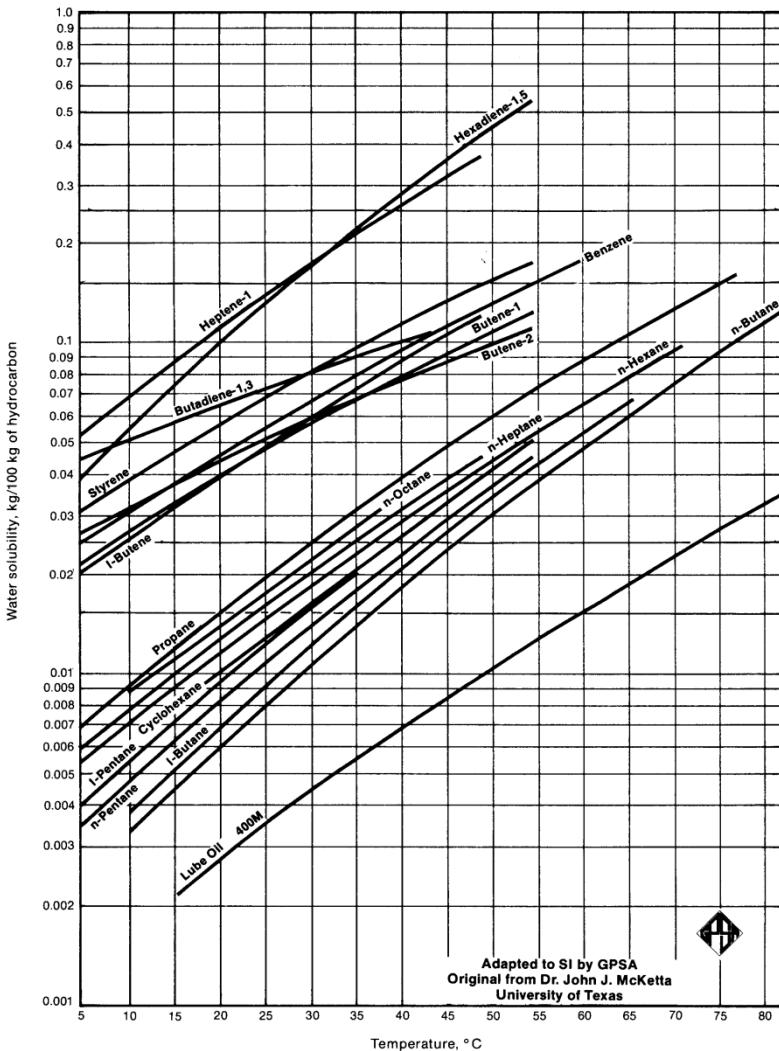
Liquid Drying

Many liquid streams must be dehydrated to allow further processing or meet requirements of a handling chain to a direct consumer. Commercial propane must be dry before entering the fuel market to prevent freezing problems as the liquid vaporizes at temperatures below the hydrate point, or even below the freezing point of any free water that may be present. The amount of water that can be in solution with light hydrocarbon liquid is very small, even at the saturation point.

Effective drying to very low levels of moisture is usually required.

The solubility of water in liquid hydrocarbons is shown. The desired maximum moisture level for commercial propane is approximately 10 mg/kg. However, liquids exposed to cryogenic temperatures require virtually all the moisture be removed.

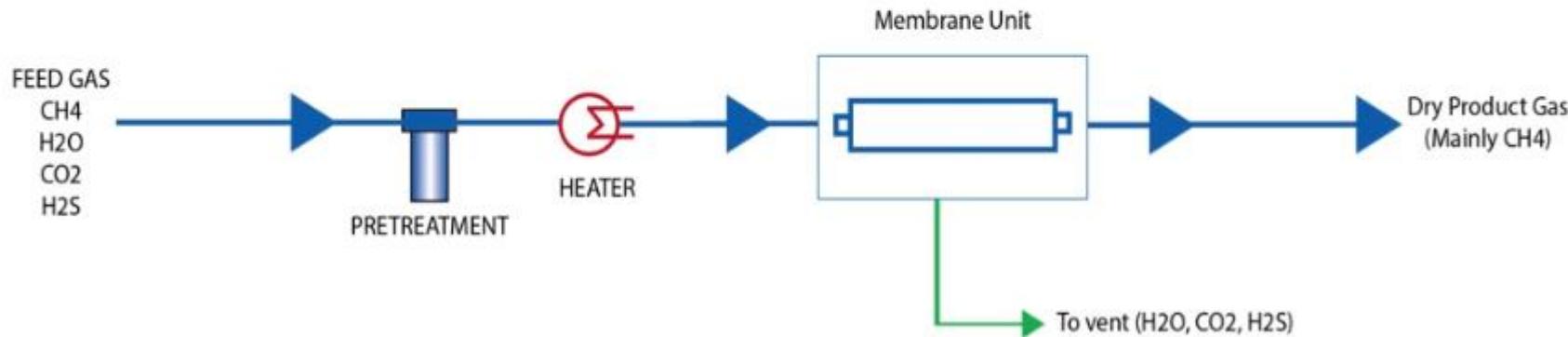
Molecular sieves are often used to dry liquid hydrocarbons from gas plants.



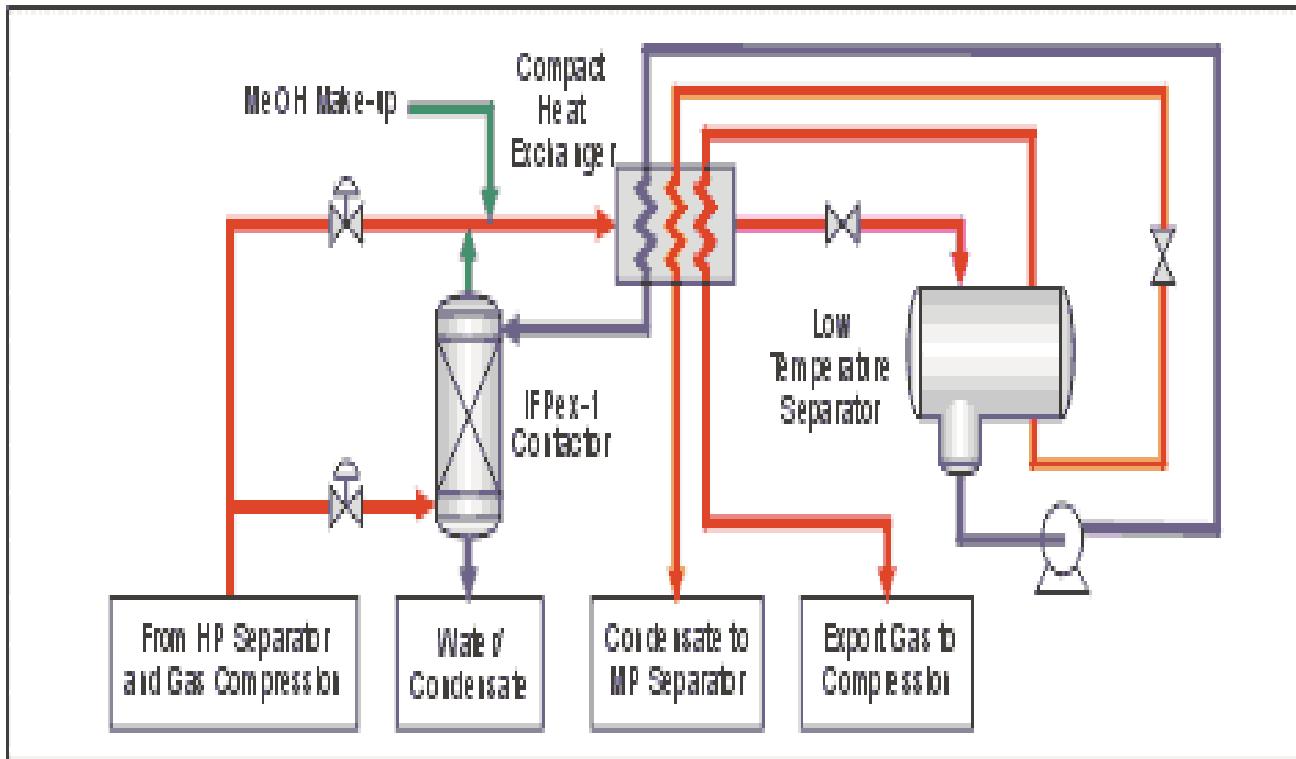
Membrane Dehydration

Membrane dehydration is yet to gain widespread acceptance. In a typical membrane gas dehydration system the feed gas is first filtered to remove any entrained liquids. The gas then enters the membrane modules. The H₂O as well as the CO₂ and H₂S permeate preferably through the membrane. The non-permeated gas, mainly CH₄, remains at pressure and is the product gas.

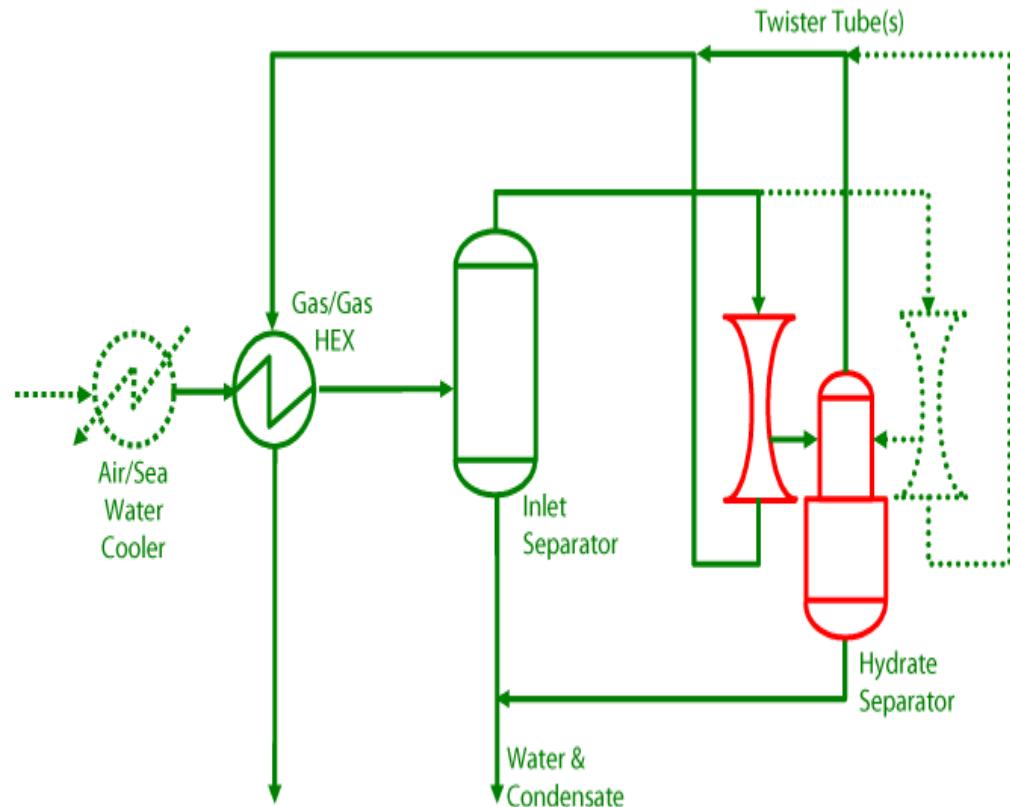
Membrane Gas Dehydration



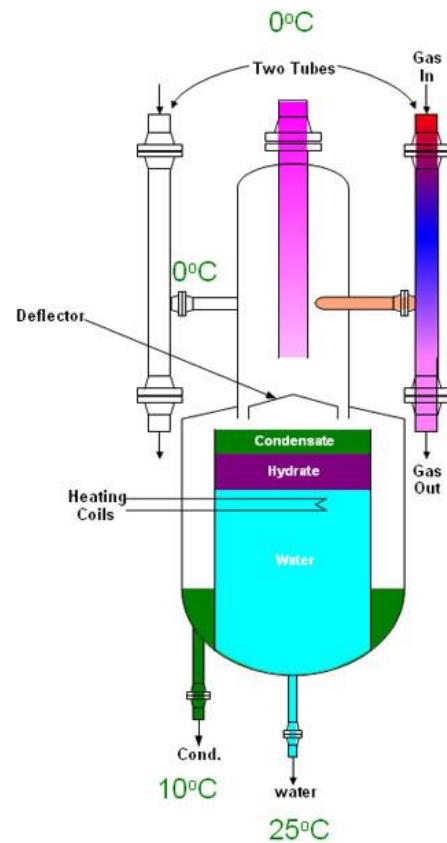
IFPEX



Shell Twister - Dehydration Technology



Shell Twister - Dehydration Technology



Shell Twister - Dehydration Technology

Potential Space saving wrt Traditional technologies



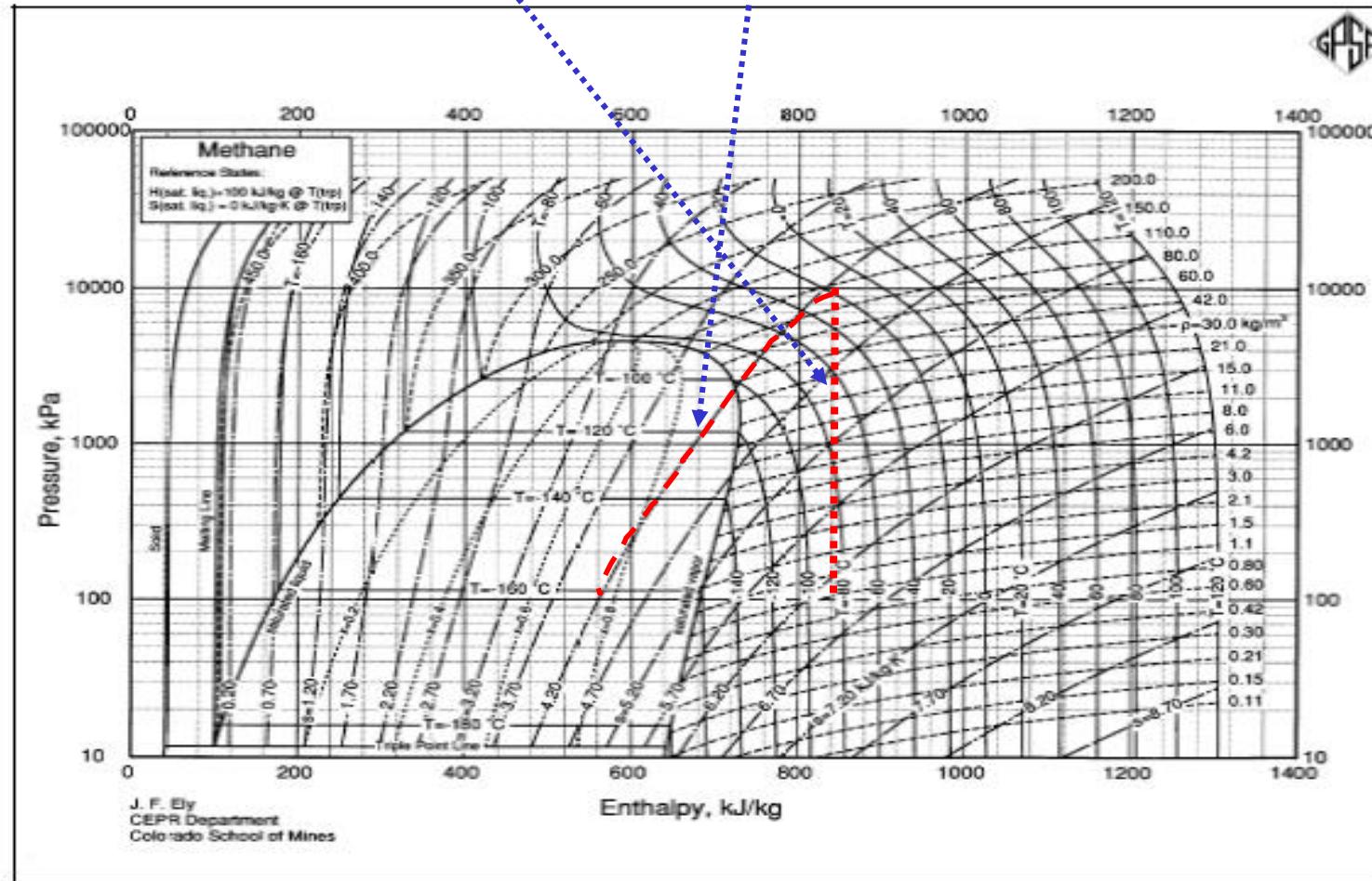
NGL Recovery/Dewpointing

In addition to achieving gas sales specification it may be commercially attractive to condense additional NGLs (mainly propane and butane) and export with oil or as a separate stream for treatment and sales.

Basic Thermodynamics

Isenthalpic/Isentropic Paths

Recap of thermodynamics which will be used for NGL treatment.



Dewpoint Control Methods

The main schemes for dewpoint control are :

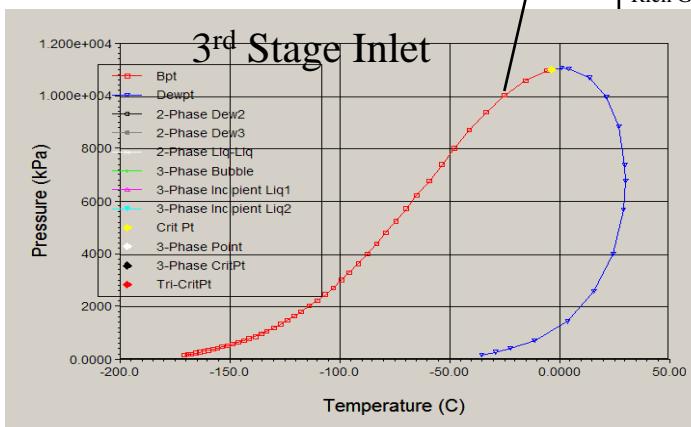
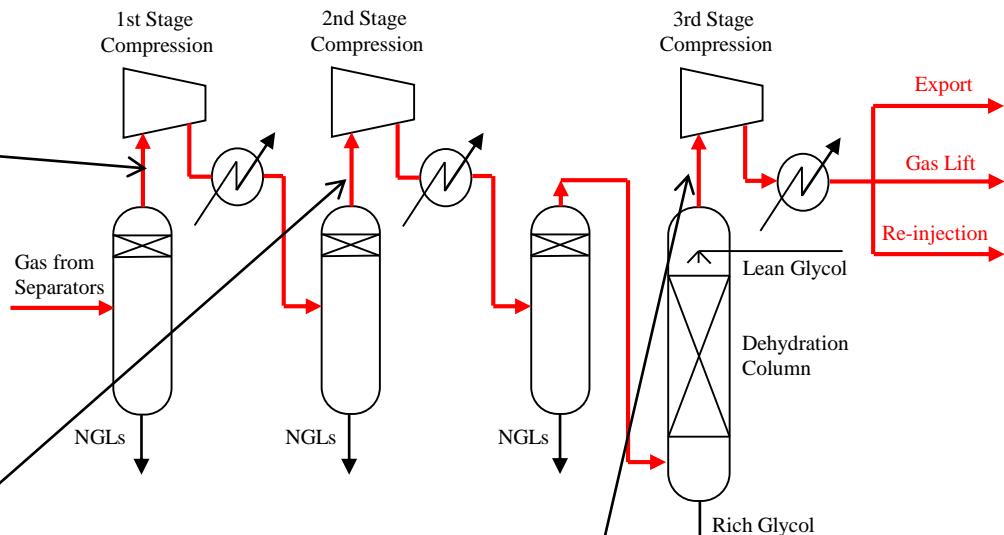
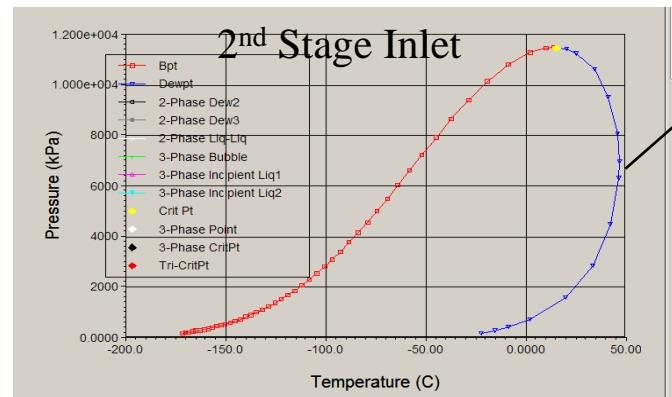
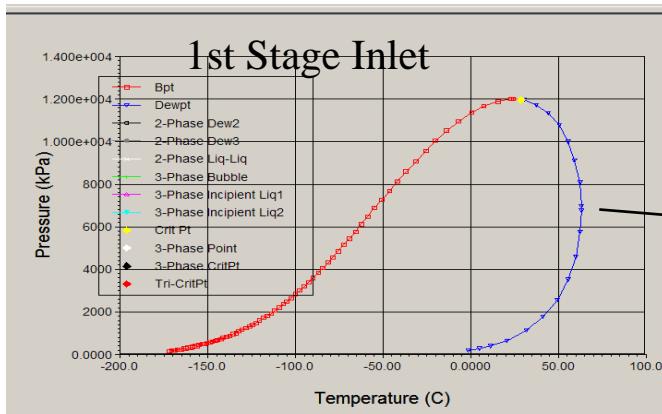
- External Refrigeration
- Joule Thomson Expansion (Adiabatic Process)
- Turbo Expander (Isentropic Process)

Each of the schemes utilise the cold gas sales and condensate streams or heat exchange to provide a more thermally efficient cycle.

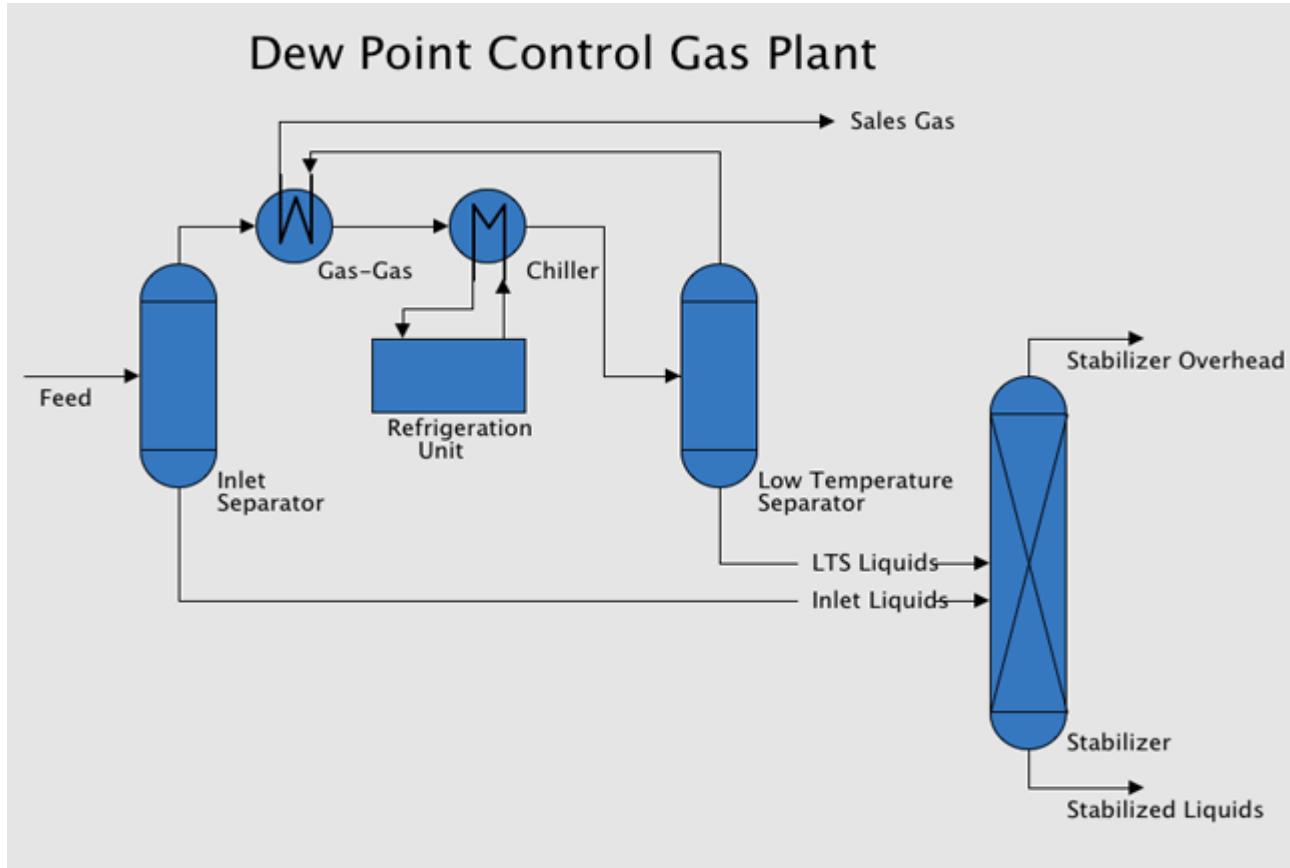
Other schemes include:-

- Silica Gel Beds
- Vortex Tubes

Gas Leaning

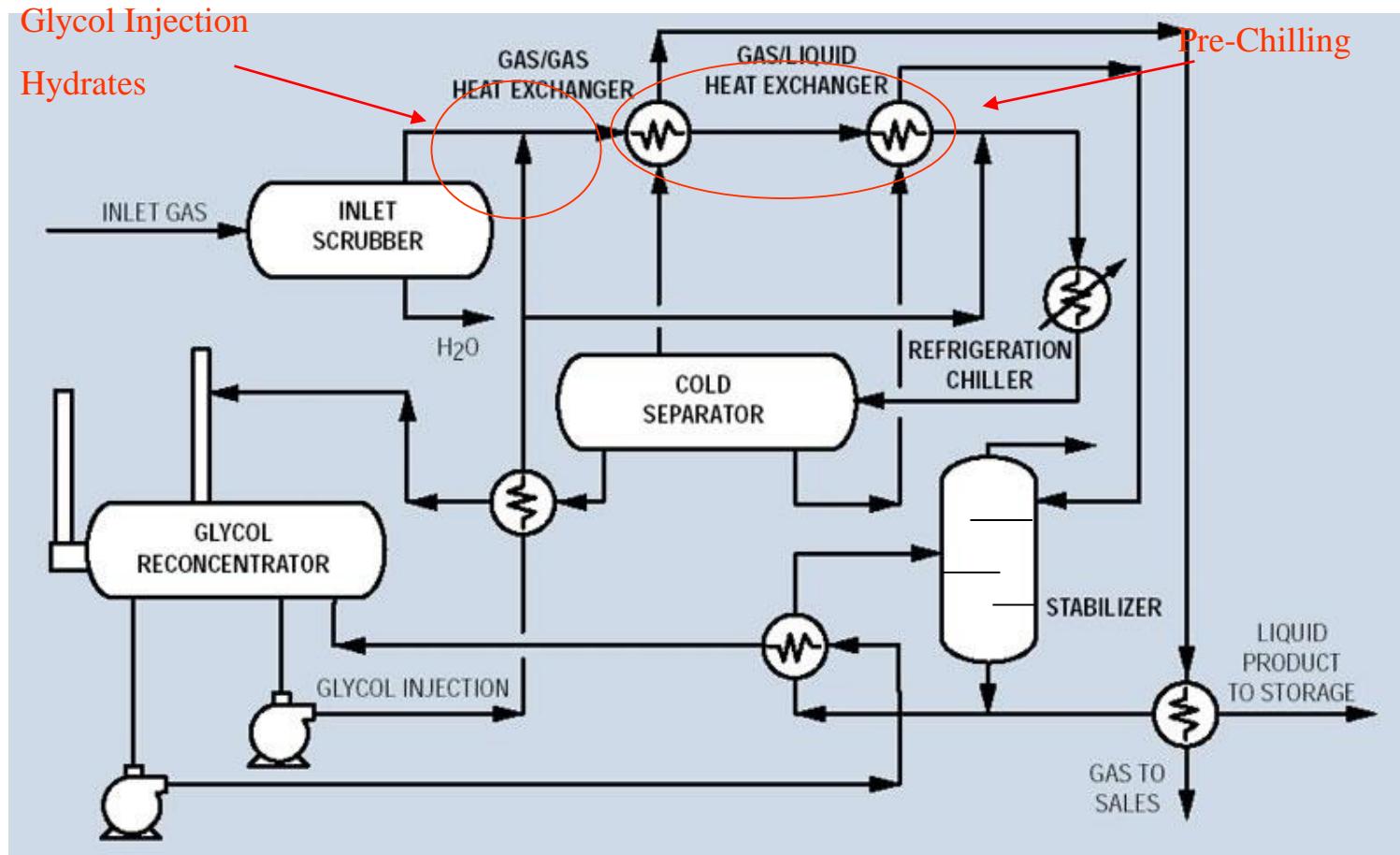


Refrigeration



The feed is cooled by external refrigeration before passing into the dewpoint separator where the heavier fractions drop out for NGL recovery leaving a low dewpoint gas.

Refrigeration PFD



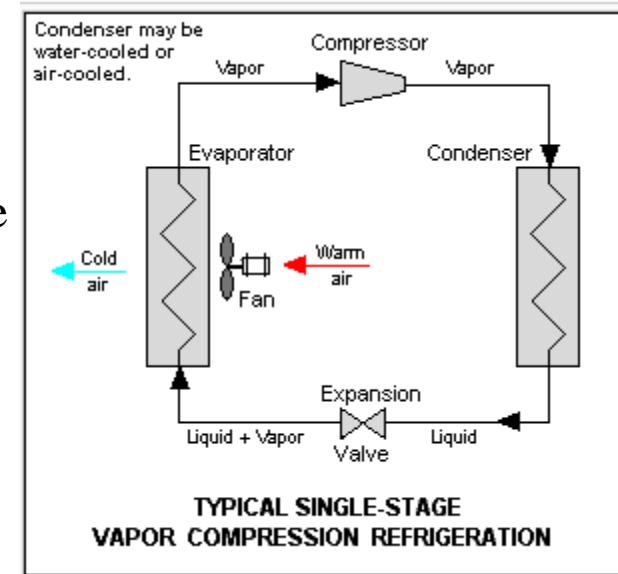
Refrigeration

The vapor-compression cycle is used in most household refrigerators as well as in many large commercial and industrial refrigeration systems. The equipment configuration is shown opposite.

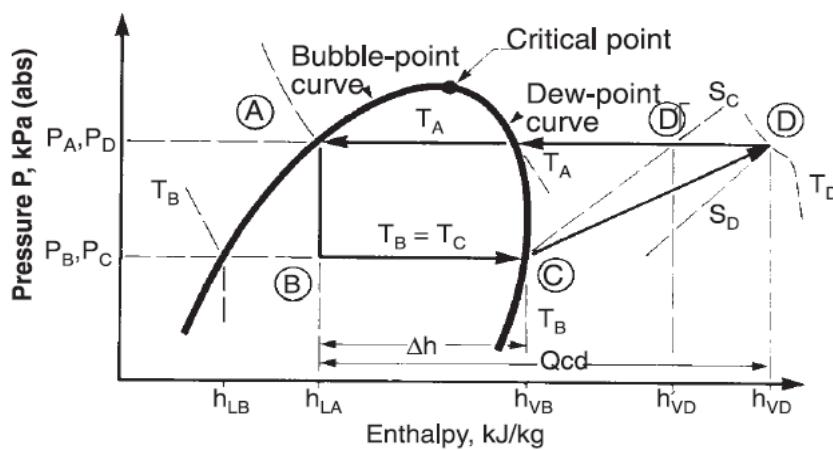
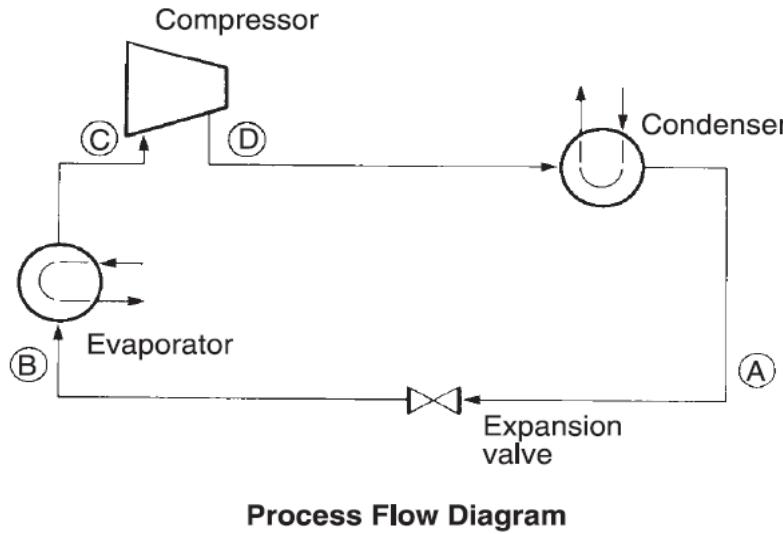
In this cycle, a circulating refrigerant such as Freon enters the compressor as a vapor. The vapor is compressed and exits the compressor superheated. The superheated vapor passes to a condenser which removes the superheat and then condenses the vapor at essentially constant pressure and temperature. The condensing medium has to be colder than the pressurised refrigerant to affect condensation. The condensing medium is often air or water.

The liquid pressurised refrigerant is expanded across a throttle valve where Joule-Thomson cooling takes place.

The cold liquid-vapor mixture is passed to an evaporator (process cooler) where it is vapourised at essentially constant temperature. The latent heat of vapourisation is the effect which cools the process. The resulting refrigerant vapor returns to the compressor inlet to complete the thermodynamic cycle.



Mechanical Refrigeration

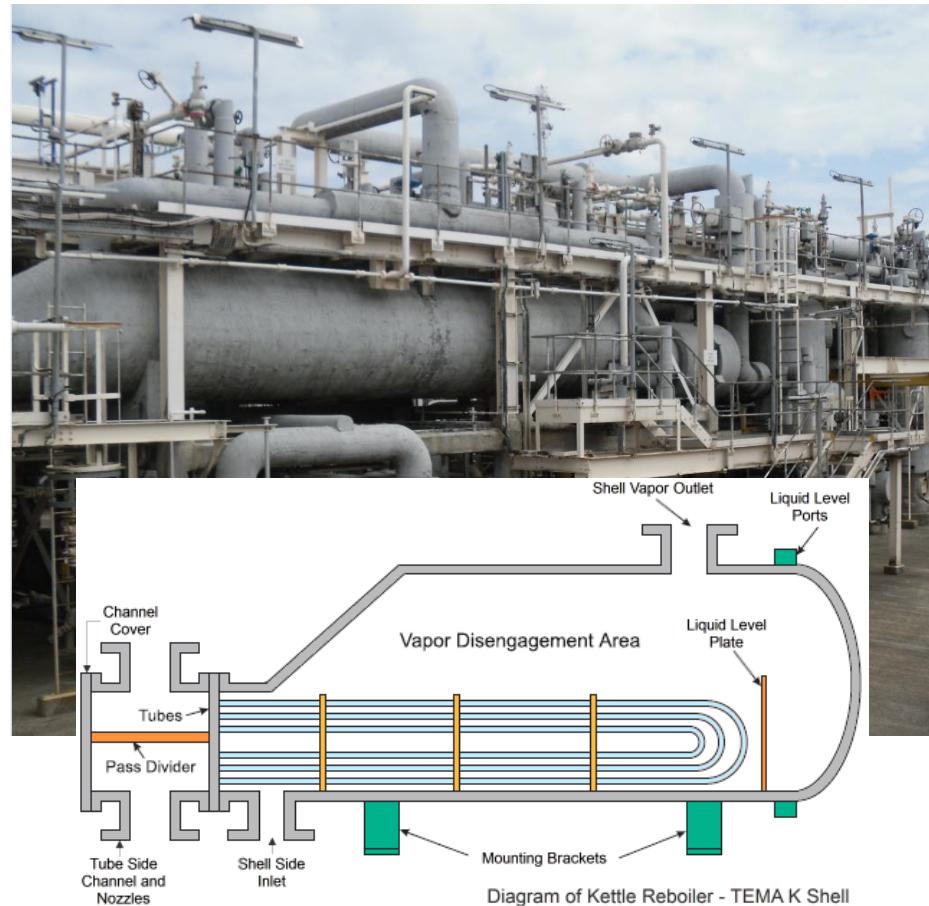


The refrigeration cycle can be broken down into four distinct steps;

- Expansion
- Evaporation
- Compression
- Condensation

Evaporators/Chillers

The most common type of chiller employed in the gas processing industry is the kettle type. The refrigerant is expanded into the shell of the kettle where a level is maintained to completely submerge the process tube bundle. When using a kettle type chiller, care should be taken to provide adequate vapour disengaging space above the operating level of liquid refrigerant. This type chiller improperly designed and operated is probably the largest single cause of compressor failure due to liquid carryover.



Refrigerants

Selection of the refrigerant is generally based upon temperature requirements, availability, safety, environment and economics. Historical refrigerants are HCFCs (Hydro-chloro-fluro-carbons) and CFCs (Chloro-fluro-carbons) of which Freon R22 is widely used. As a result of ozone depletion effects these chemical are being replaced with non-ozone depleting refrigerants. As a Party to the Montreal Protocol, countries must incrementally decrease HCFC consumption and production, culminating in a complete HCFC phase out in 2030. The major milestones that are upcoming for developed countries are a reduction in 2015 to at least 90 percent below baseline HCFC levels and a reduction in 2020 to at least 99.5 percent below baseline.

Classes of Refrigerants

Hydrocarbons and Inorganic Compounds These include ammonia, CO₂, simple hydrocarbons, and water; they are often referred to as natural refrigerants. They have zero ODP and low GWP values.

Halocarbon Refrigerants The halocarbon refrigerants include one or more of the halogens (i.e., the elements fluorine, chlorine, or much less frequently, bromine or iodine) in a molecule with a carbon backbone. These chemicals were first commercialized in the 1930s and include CFCs (i.e., containing carbon, fluorine, and chlorine), HCFCs (also containing hydrogen), and HFCs (which do not contain chlorine). The most commonly used CFCs used CFC refrigerants were CFC-12 and CFC-11. The production of these refrigerants was phased out by year end 1995 in developed countries and by year end 2009 in developing countries due to their impact on stratospheric ozone.

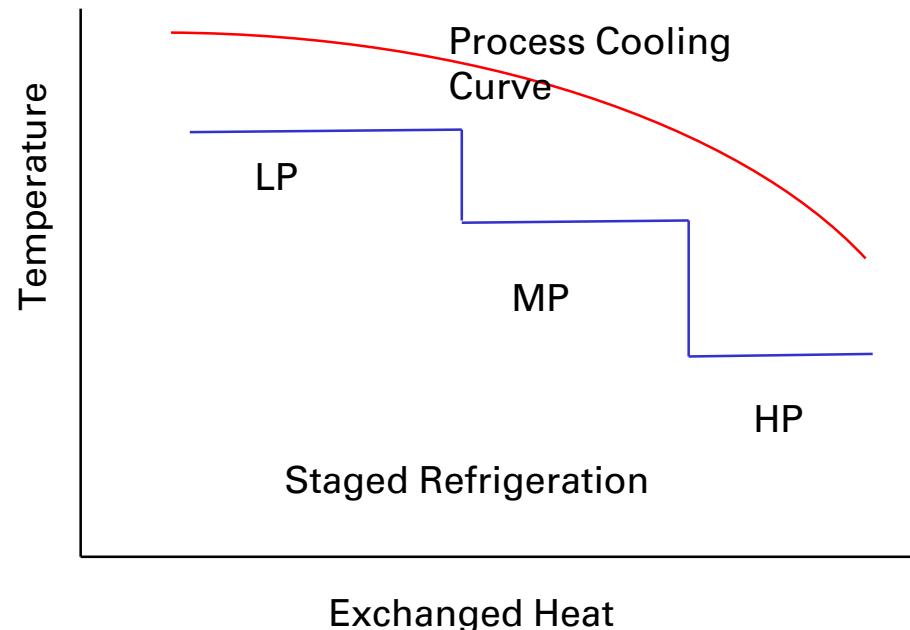
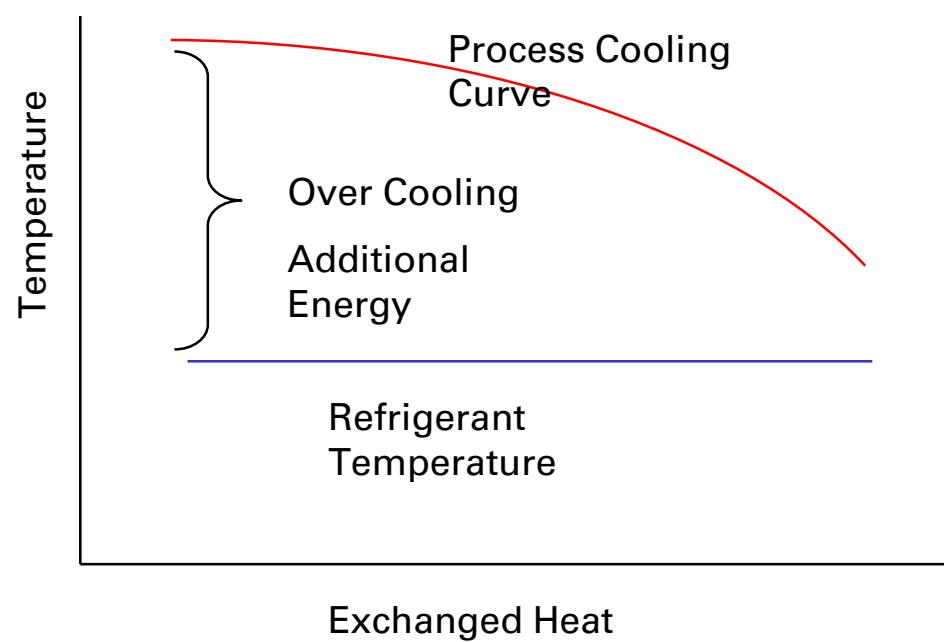
Refrigerant Summary

Refrigerant	GWP	ODP	Comments
CFC 12	8100	1	Banned in EU since 2000.
HCFC 22	1500	0.05	Being phased out. Use of these refrigerants should be identified and plans for their replacement developed in accordance with dates set out in regional legislation.
HFC 134a	1300	0	Various HFCs used since mid-1990s as alternatives to ozone-depleting CFCs and HCFCs in a wide variety of stationary and mobile RAC applications. Three examples given here – around 20 others also available. HFC 134a is being phased out in the EU and is proposed to be phased out in the US for mobile air conditioning systems.
HFC 404A	3300	0	High global warming potential and subject to regulation in the EU, with likely regulation in the US. Potential for phase out in the longer term as alternatives become available.
HFC 410A	1725	0	
FC	~10	0	New refrigerants; very low GWP. Not yet commercially available.
CO ₂	1	0	Operates at very high pressure.
Hydrocarbons	3	0	Widely used in very small systems; highly flammable.
Ammonia	0	0	Used in large industrial systems; toxic and combustible.

ODP – ozone depletion potential

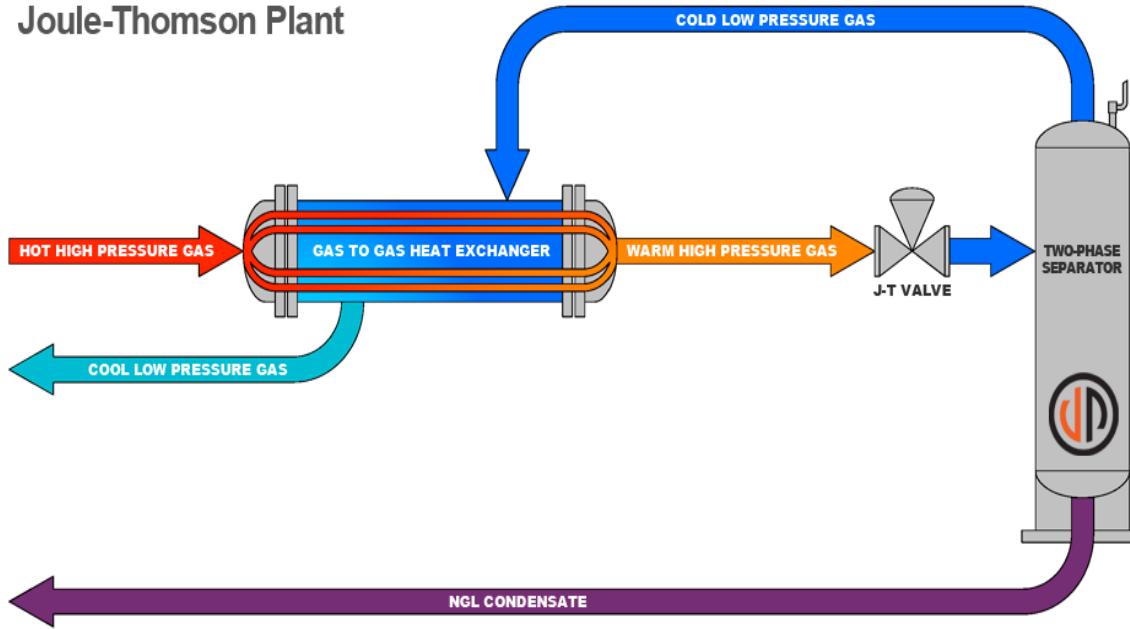
GWP – global warming potential

Staged Refrigeration



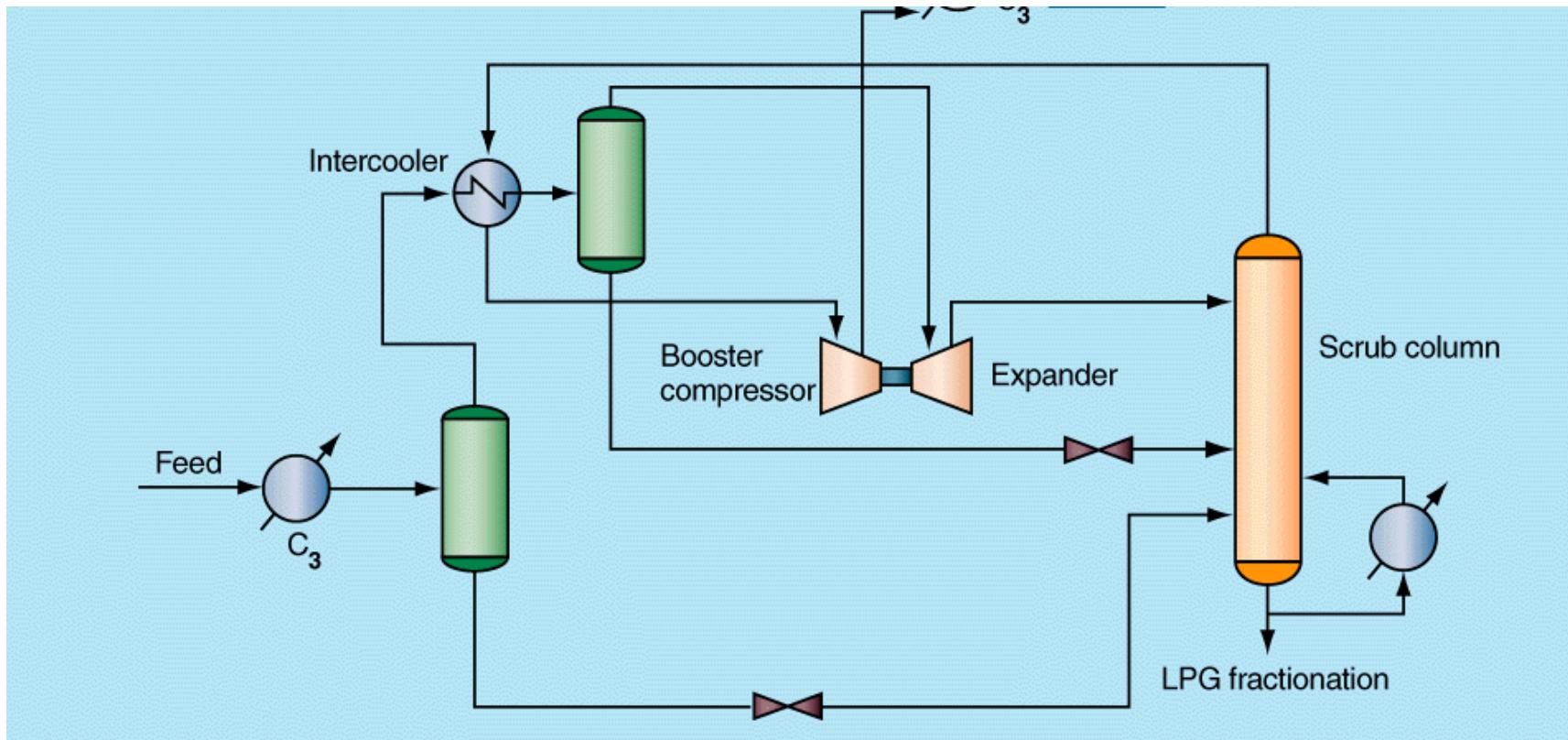
Joule Thomson Valve

Joule-Thomson Plant



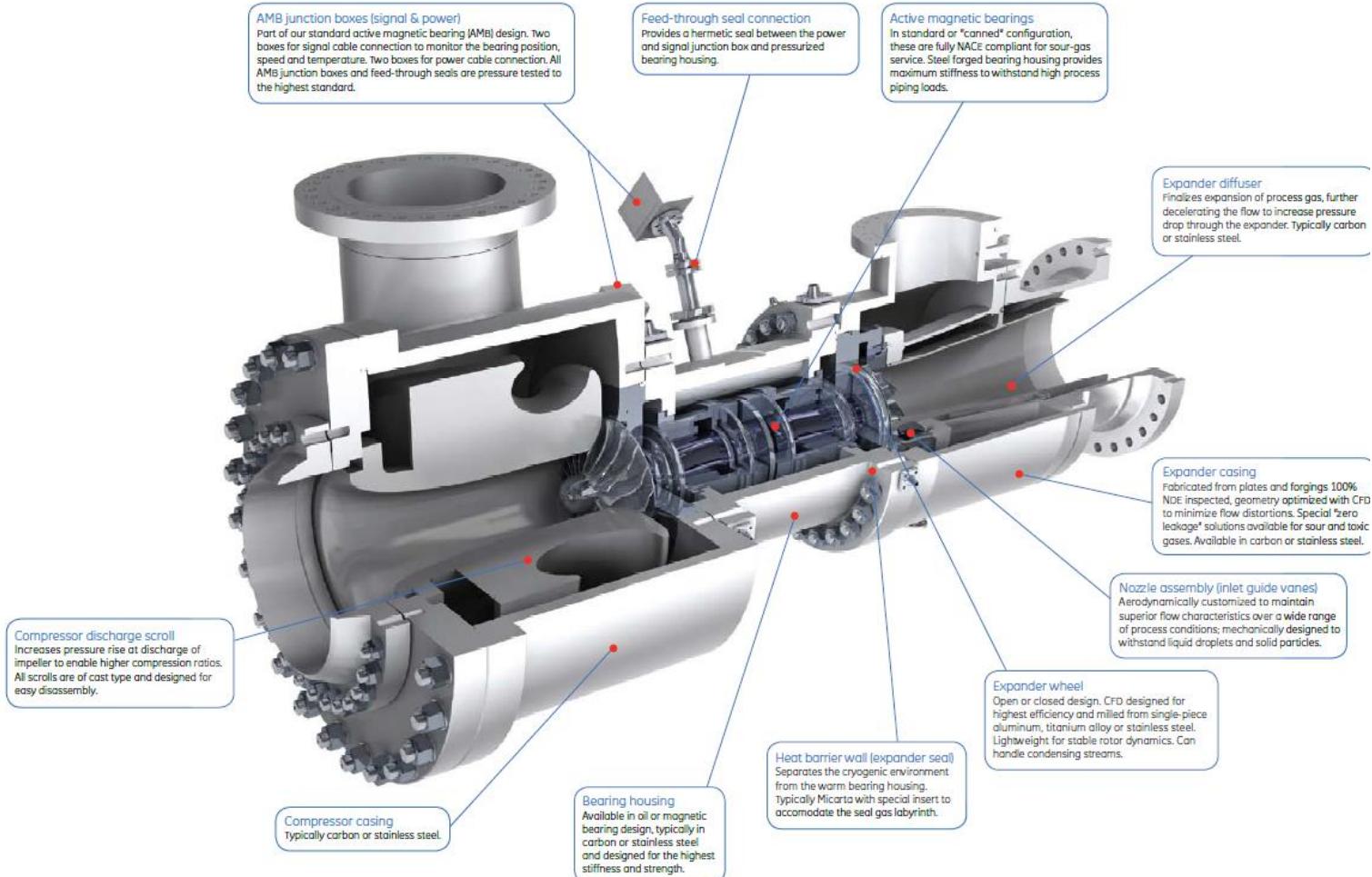
The feed is first cooled and then expansion is achieved by pressure let-down through a Joule-Thomson valve. The liquid is taken off in the separator drum.

Turbo Expansion

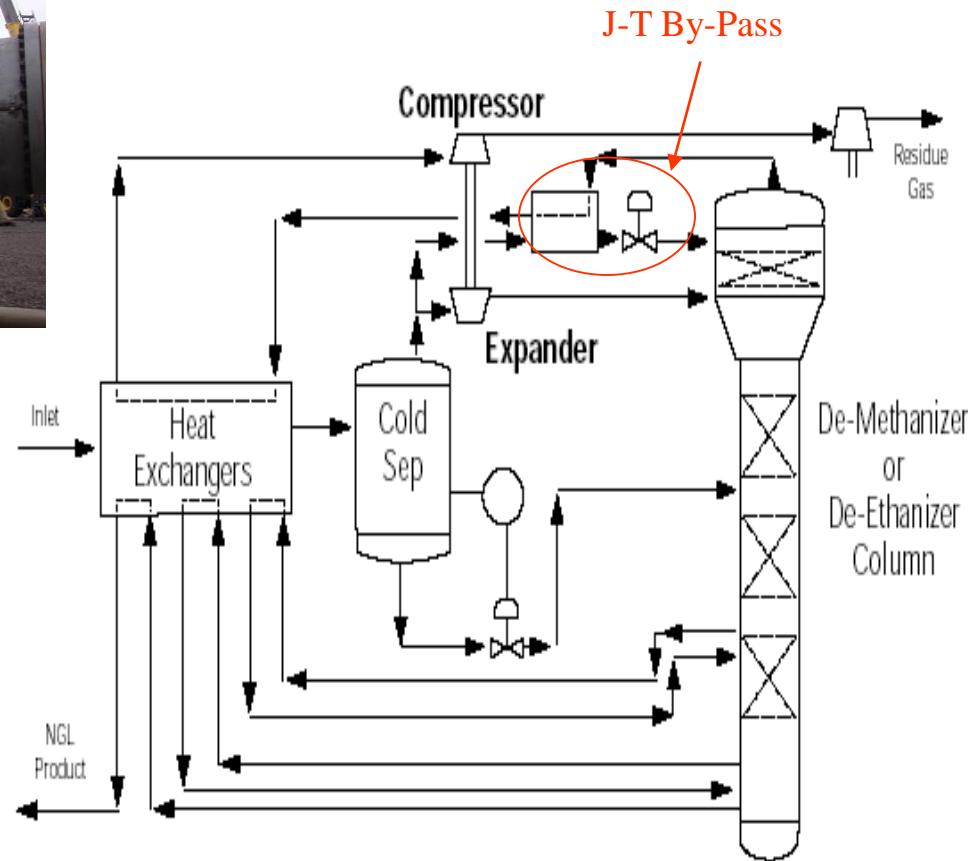


The feed is expanded through an expansion wheel, where the gas cools through pressure let down and the extraction of energy. The liquid is separated and the gas is recompressed by a second compression wheel using the power generated from the original expansion.

Turbo Expansion



Turbo Expansion



Design Guidelines

Design Guidelines

Heat exchanger approach temperature can be as follows :

- Gas / Gas exchanger approach between hot and cold streams 5 - 10°C,
- Refrigeration cooler approach between cold gas and refrigerant streams 5 - 10°C.

These temperature approaches represent initial estimates and should be subject to optimisation as design is developed. For a turbo expander / recompressor scheme typical polytropic efficiencies are :

- Turbo Expander 75%
- Recompressor 70%

Other criteria for the turbo expander selection are :

- maximum proven machine approx. 5 MW (Check with vendors for latest status)
- maximum pressure ratio 2.5 : 1
- liquid content of outlet up to 40% wt

Cold Temperature Option Selection

Deciding whether to adopt a refrigeration unit, a turbo-expander or a Joule-Thomson valve will involve a multi criteria assessment. When faced with alternative options to provide the same goal, in the case low temperature, the engineer must balance the following areas;

Capital Cost – CAPEX – how much does it cost to buy and install?

Operating Cost – OPEX how much does it cost to operate?

Availability – plant uptime, frequent breakdowns will impact production

Revenue – how much cash flow will the option deliver?

Safety – what are the key safety implications of the competing options?

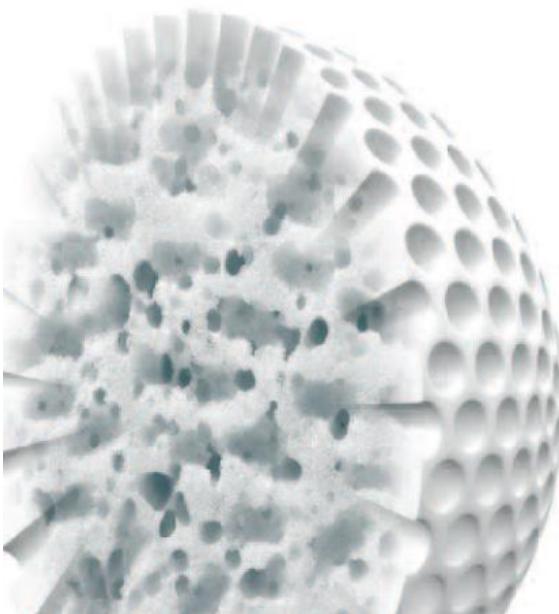
Environmental – what is the environmental footprint of the competing options?

Silica Gel

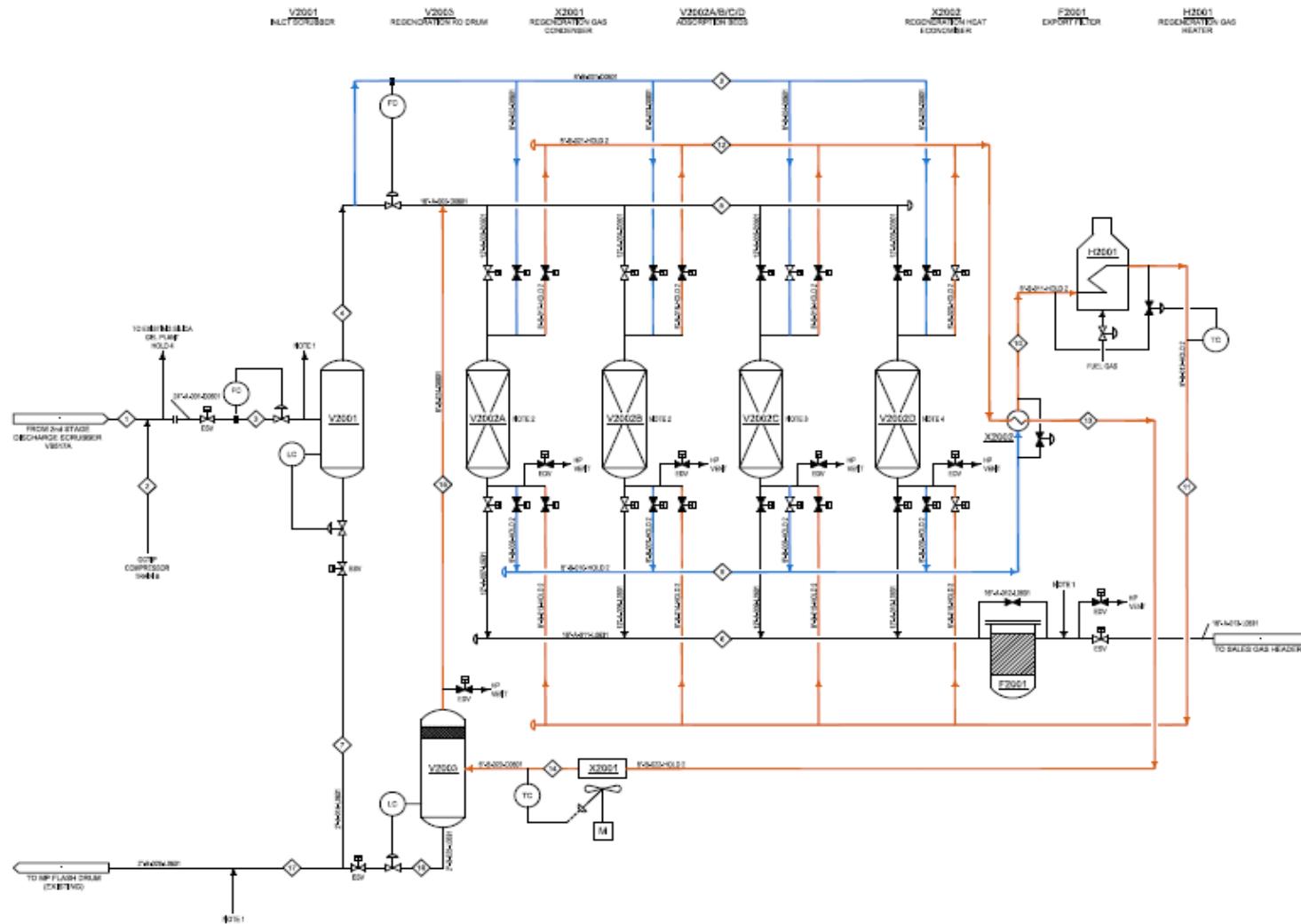
Silica gel is a porous, amorphous form of silica (SiO_2). Although it has the same chemical composition as sand, silica gel is radically different from other SiO_2 -based materials, due to its unique internal structure.

The internal structure of silica gel is composed of a vast network of inter-connected microscopic pores that attract and hold water, alcohols, hydrocarbons and other chemicals on their surface through phenomena known as physical adsorption and capillary condensation. This phenomena can be reversed by changing the conditions under which adsorption takes place, a process called regeneration.

In some circumstances silica gels can be used to adsorb the heavier hydrocarbon components in a gas to meet transportation or treatment specifications.

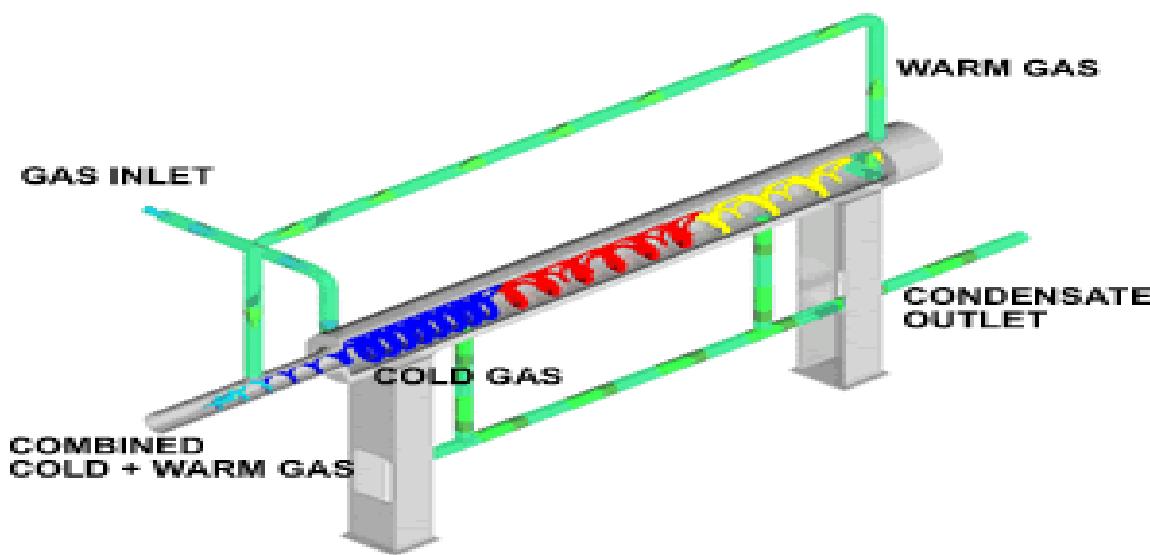


Silica Gel



Vortex Tube

The Hilsch effect: This refers to the tangential introduction of a gas stream through a nozzle creating a forced vortex which leads to a simultaneous separation of the gas and liquid in the gravitational field of the vortex and at the same time causing a temperature gradient to form across the vortex (the temperature reduces from the wall to the tube centre). Due to the temperature gradient the expansion moves away from a Joule Thomson process (adiabatic expansion) towards a more efficient isentropic expansion.



Acid Gas Removal

Hydrogen Sulphide (H₂S) and Carbon Dioxide (CO₂) are often found with oil and gas fields. The requirement for acid gas treatment is dependant on various criteria.

H₂S Removal

- Acid gas treatment may be required to limit the concentration of H₂S in the fuel gas, pipeline and sales specifications.
- Minimising impact on environment

CO₂ Removal

- Where economics dictate the requirements i.e. where savings in compression and transportation costs outweigh the cost of the treatment unit;
- For the use of existing pipeline;
- To avoid carbonic corrosion in wet pipelines
- For enhanced oil recovery (EOR) schemes
- Minimising environmental emissions
- Carbon sequestration

Other components which may require treatment/consideration are;

- Hydrogen Cyanide (HCN)
- Carbonyl Sulphide (COS)
- Carbon Disulphide (CS₂)
- Mercaptans (RSH)
- Nitrogen (N₂)
- Sulphur Dioxide (SO₂)
- Mercury (Hg)

Hydrogen Sulphide Removal

Health Issues - H₂S emissions from oil & gas development have the potential to pose a significant health risk and is odourless at concentrations above 0.2ppm because it quickly degrades the sense of smell.

H ₂ S Concentration (ppm)	Health Issue
≤ 0.2	Odour threshold
10	Eye irritation
100	Headache, dizziness, vomiting, heavy coughing
200 – 300	Eye & respiratory tract inflammation
500 – 700	Loss of consciousness – <i>death within 30 – 60 minutes</i>
700 – 900	Rapid loss of consciousness – <i>death within 30 minutes</i>
≥ 1000	Unconsciousness in seconds – <i>death within minutes</i>

Flammability Issues - H₂S is an extremely flammable substance - H₂S vapour can travel considerable distance to a source of ignition and flash back explosively.

CO_2 and H_2S removal using Amine Gas Treating

There are many different amines used in gas treating:

Monoethanolamine (MEA)

Diethanolamine (DEA)

Methyldiethanolamine (MDEA)

Diisopropylamine (DIPA)

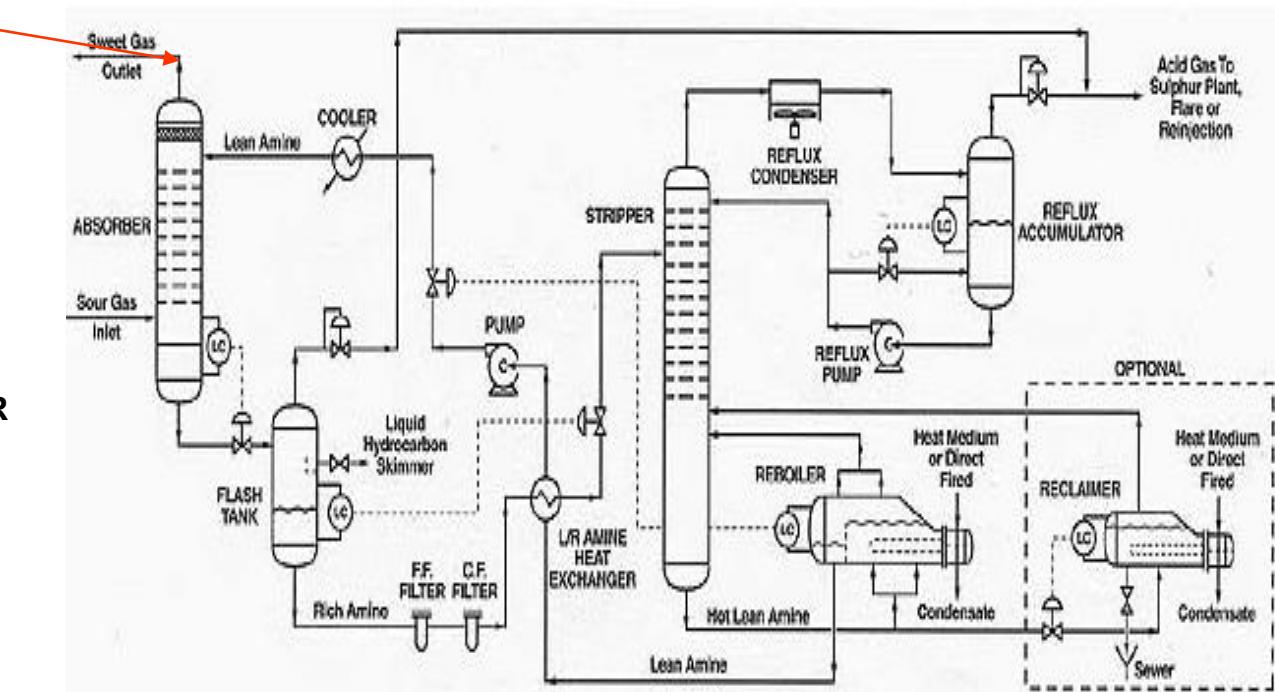
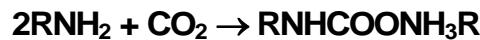
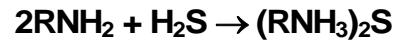
Aminoethoxyethanol (diglycolamine) (DGA)

The most commonly used amines in industrial plants are the alkanolamines MEA, DEA, and MDEA. The chemical suppliers often increase the reaction efficiency by including proprietary activators with the amine.

Amine Based Chemical Solvent Processes

Water Saturated

Basic Amine
Sweetening
Plant



Typical Properties /
Process Conditions

Comp	P _{vap} @ 38 C mbar	Aq. Sol. wt%	T cont C	T strip C
MEA	1.400	10 - 20	30 - 60	90 - 115
DEA	0.077	10 - 20	30 - 60	90 - 115
DGA	0.213	40 - 70	60 - 85	125 - 140

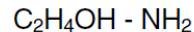
Chemical Solvent Processes

Amine Solvent Chemistry

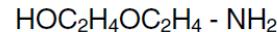
There are many amines available and the particular choice of amine will depend on the composition of the feed gas and specification required. Amines can be categorised by the number of organic groups bonded to the central nitrogen atom, as primary, secondary or tertiary. Examples are listed below:

PRIMARY AMINES

Monoethanolamine (MEA)

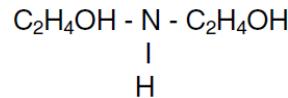


DIGLYCOLAMINE Agent (DGA)

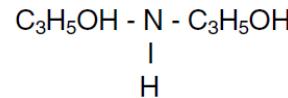


SECONDARY AMINES

Diethanolamine (DEA)

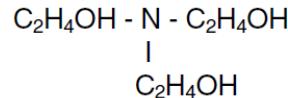


Diisopropanolamine (DIPA)

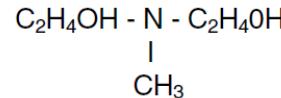


TERTIARY AMINES

Triethanolamine (TEA)



Methyldiethanolamine (MDEA)



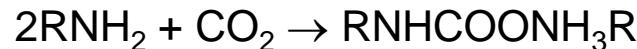
Chemical Solvent Processes

Amines with stronger base properties will be more reactive towards H₂S and CO₂ and will form stronger chemical bonds. As the amine reactivity increases, i.e. from tertiary amine to primary amine, then the acid gas vapour pressure over the amine will be lower for a given loading and a higher equilibrium loading may be achieved.

The reactions between an amine and H₂S and CO₂ occur according to the following equations (shown for a primary amine):



CO₂ reacts predominantly via the reaction below, forming a carbamate with primary and secondary amines.

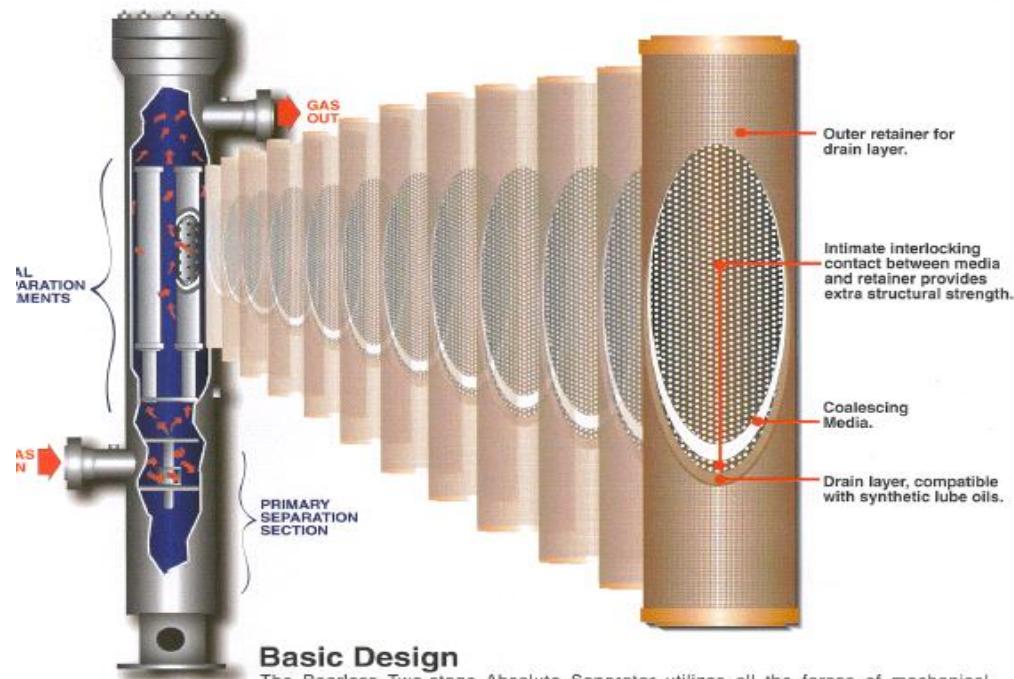


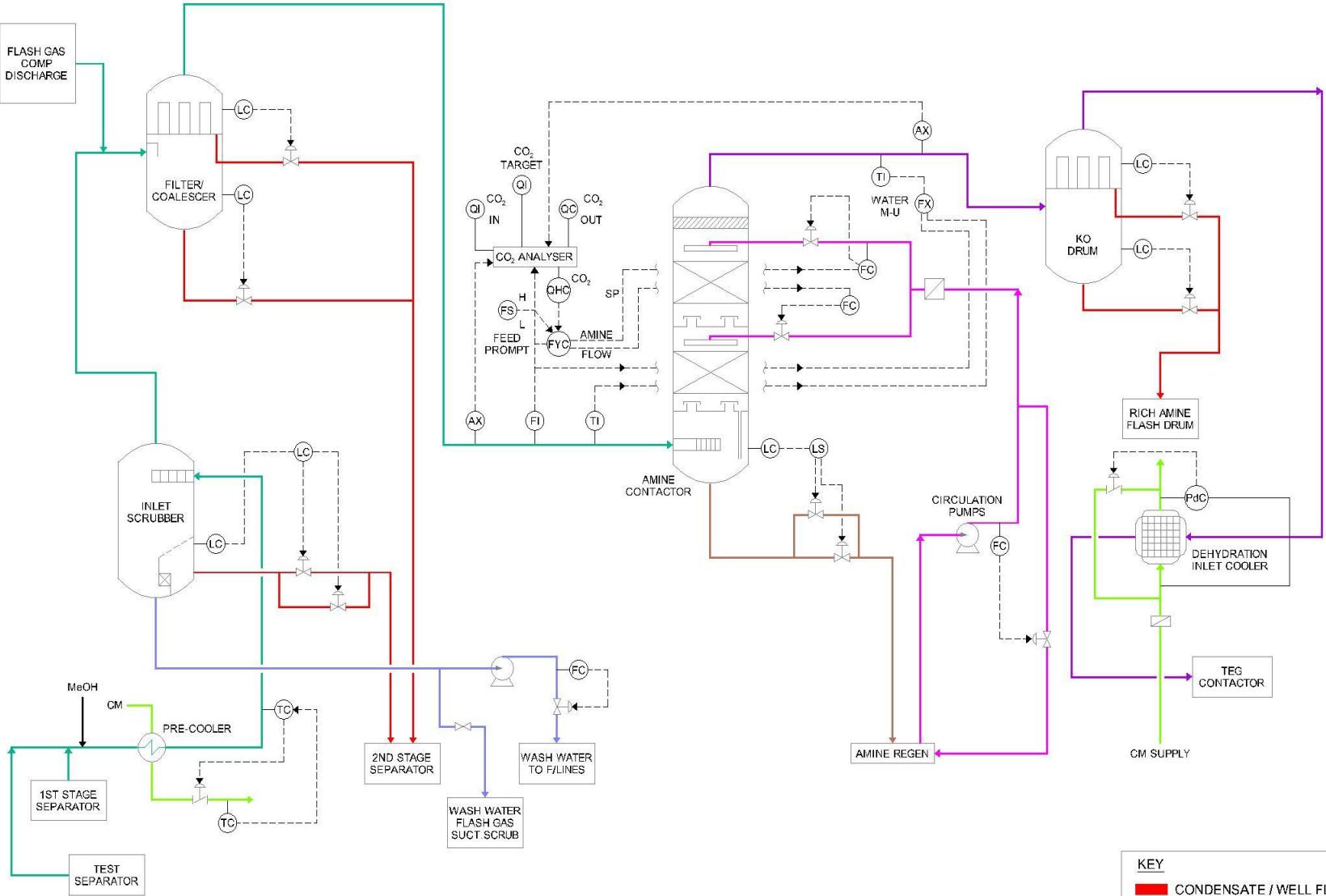
The main difference with tertiary amines compared to primary and secondary amines is that CO₂ does not react to form carbamate.

Inlet Filter Coalescer

Coalescing elements with a controlled pore structure are used to filter and coalesce entrained liquid.

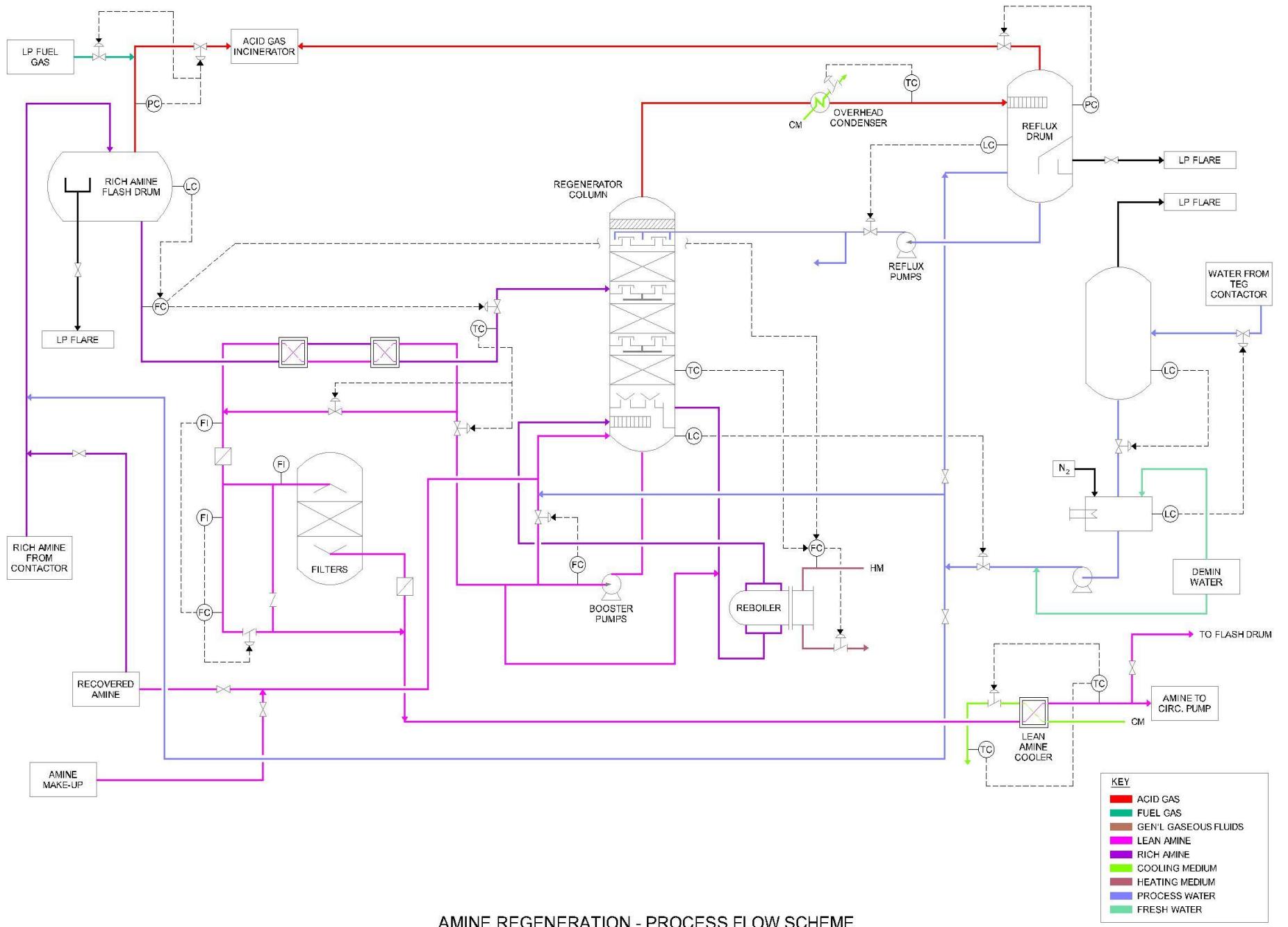
Droplet capture is around 0.3-0.5 micron.





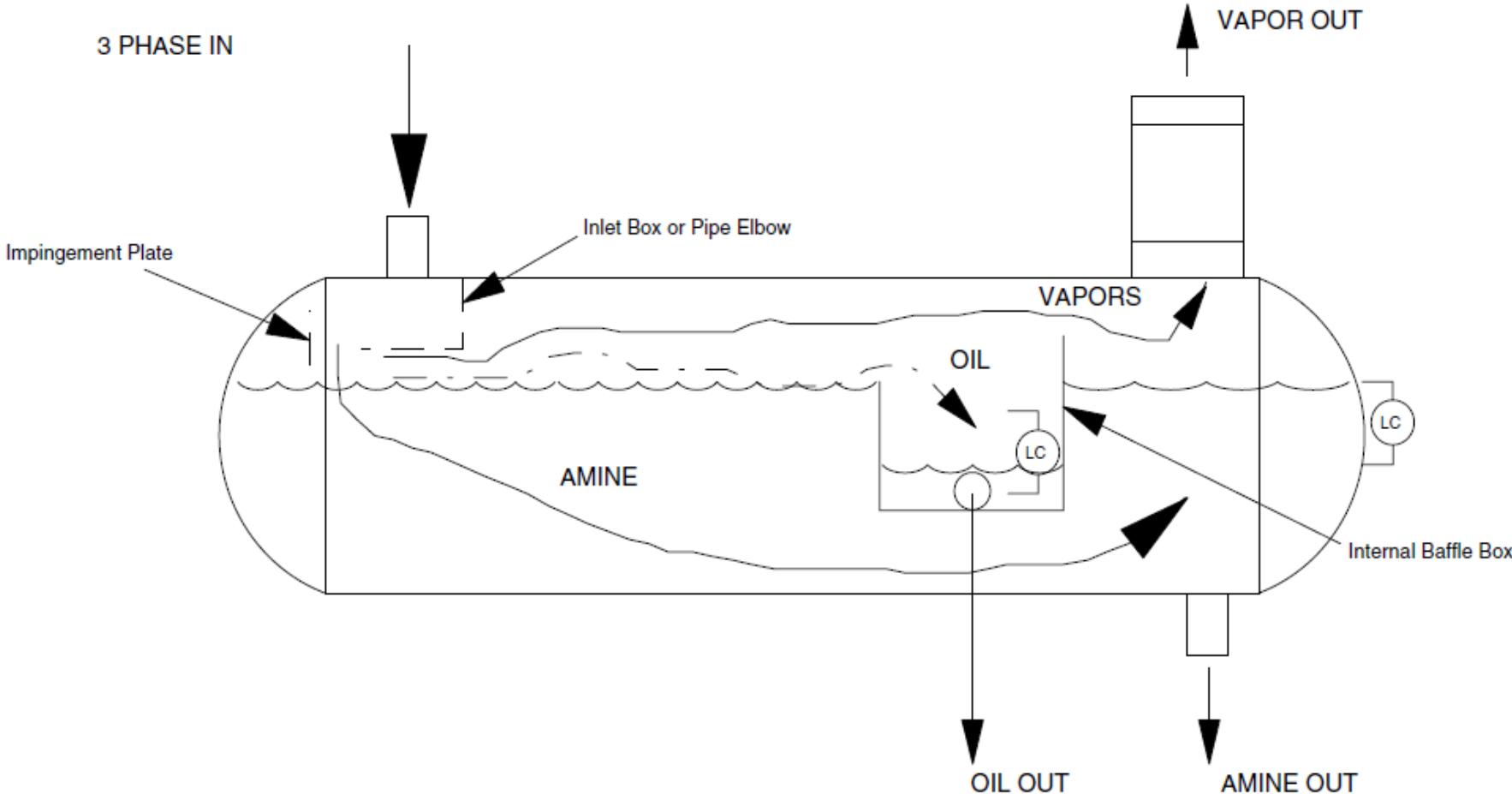
AMINE SWEETENING - PROCESS FLOW SCHEME

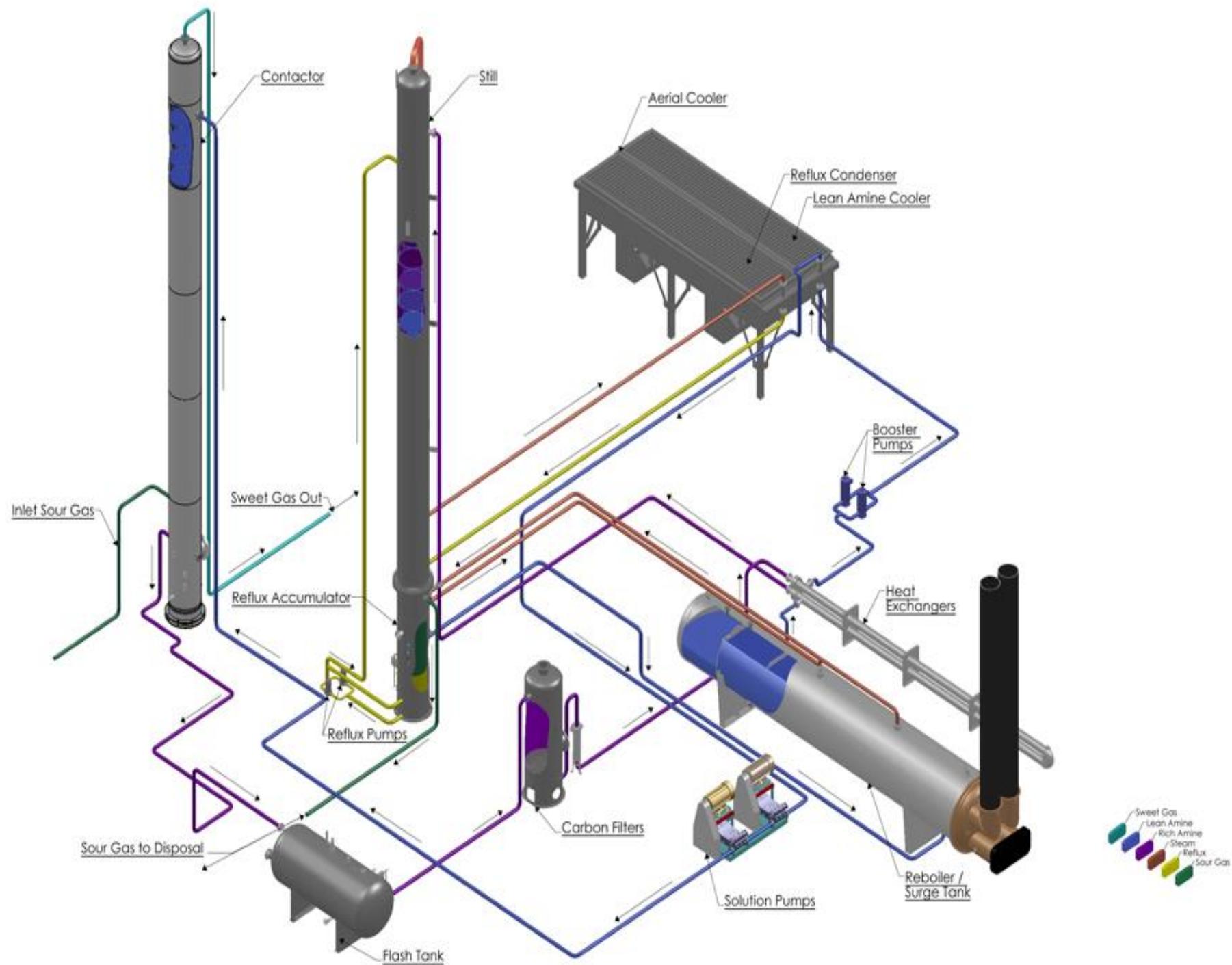
KEY
CONDENSATE / WELL FLUID
WET SOUR GAS
WET SWEET GAS
LEAN AMINE
RICH AMINE
COOLING MEDIUM
PROCESS WATER



AMINE REGENERATION - PROCESS FLOW SCHEME

Typical Flash Drum Arrangement





Amine Contaminants – Operational Issues

AMINE CONTAMINANTS

Amine contaminants can be grouped into five distinct categories; (1) heat stable salts, (2), degradation, (3) injection chemicals, (4) hydrocarbons and (5) particulates. All of these contaminant categories can typically be present in any given amine system at the same time, although the amount of each one can vary from insignificant to several per cent.

Physical Solvent Processes

These processes are based on physical absorption. In general, the physical solvent process should be considered when :

- The partial pressure of acid gas is above 50 psi
- The heavy hydrocarbon concentration in the feed gas is low
- Bulk removal of the acid gas is desired
- Selective removal of H_2S is desired

Physical Solvent Processes

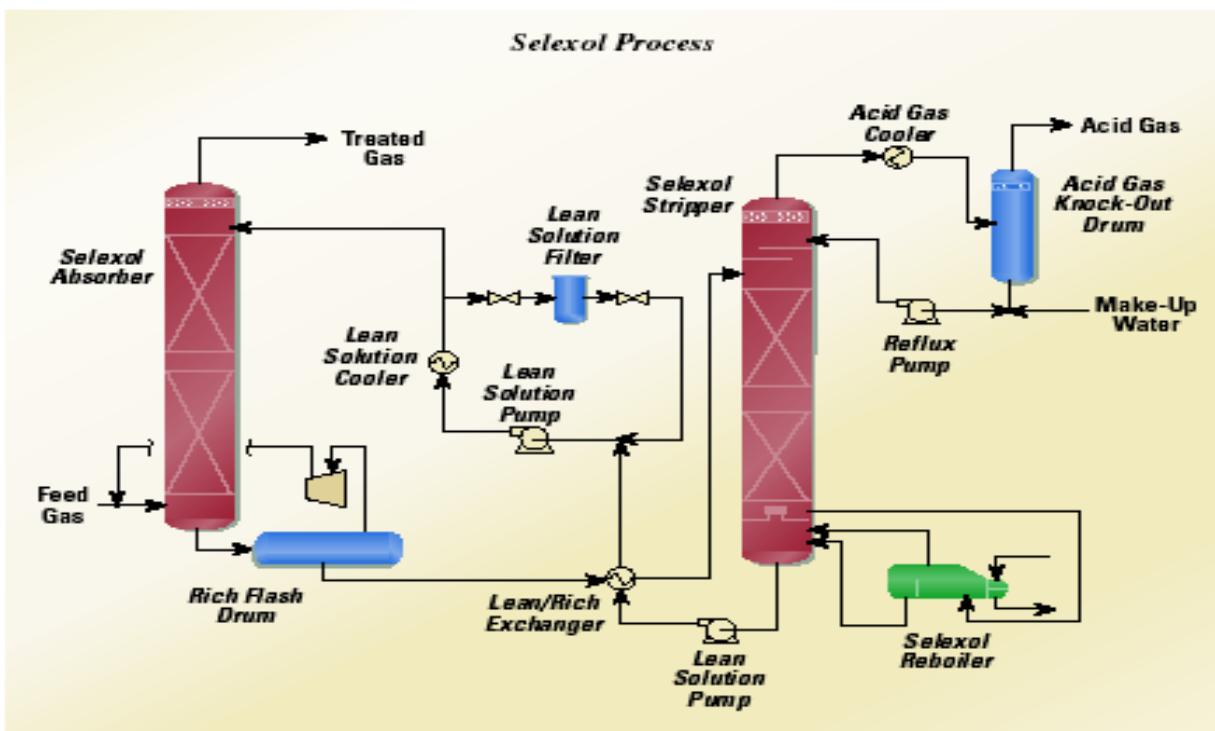
Processes are characterised by :

- Low heats of solution
- Low utility requirements
- Co-absorption of heavy hydrocarbons (C₃+) from the gas stream
- Sensitivity of treated gas specification and solvent circulation rate to the absorption pressure

Generally physical solvent processes show advantages over chemical solvent processes at high absorption pressures. The solution loading and treated gas specification are directly proportional to the acid gas partial pressure in the feed gas.

Physical Solvent Processes

Typical Gas Sweetening by Physical Adsorption



- Process operates at ambient / sub-ambient temperatures to enhance the solubility of the acid gases
- Solvents are relatively non-corrosive so low cost materials can be used.
- Chemical losses are low due to low solvent vapour pressure.

Molecular Sieves

Pore size	Type of Molecular Sieve	Application
3A	Potassium variation of sodium alumina silicate	Commercial dehydration of unsaturated hydrocarbon streams, including cracked gas, propylene, butadiene, acetylene; drying polar liquids such as methanol and ethanol. Adsorption of molecules such as NH ₃ and H ₂ O from a N ₂ /H ₂ flow. Considered a general-purpose drying agent in polar and nonpolar media.
4A	Sodium alumina silicate	Preferred for static dehydration in closed liquid or gas systems, e.g., in packaging of drugs, electric components and perishable chemicals; water scavenging in printing and plastics systems and drying saturated hydrocarbon streams. Adsorbed species include SO ₂ , CO ₂ , H ₂ S, C ₂ H ₄ , C ₂ H ₆ , and C ₃ H ₆ . Generally considered a universal drying agent in polar and nonpolar media
5A	Calcium variation of sodium alumina silicate	Separation of normal paraffins from branched-chain and cyclic hydrocarbons; removal of H ₂ S, CO ₂ and mercaptans from natural gas. Molecules adsorbed include nC ₄ H ₁₀ , nC ₄ H ₉ OH, C ₃ H ₈ to C ₂₂ H ₄₆ , and dichlorodifluoro-methane (Freon 12 [®]).
13X	Sodium alumina silicate with X crystal structure	Commercial gas drying, air plant feed purification (simultaneous H ₂ O and CO ₂ removal) and liquid hydrocarbon/natural gas sweetening (H ₂ S and mercaptan removal).

Molecular Sieves

The adsorption of components on the sieve is:

- Directly proportional to molecular weight.
- Inversely proportional to vapour pressure.
- Directly proportional to dipole moment (polarity).

Adsorption in order of retention:

- Water (strongest).
- Methanol (co-adsorbs with water).
- Heavier mercaptans.
- Light mercaptans H_2S .
- CO_2 (similar to H_2S).
- COS (weakest).

A common use of molecular sieves is mercaptan removal. Molecular sieves are also used for H_2S removal to about 1000 ppm.

Plants may be designed to adsorb H_2S and water whilst leaving CO_2 in the gas stream.

Batch Solvents/Beds

Sulfatreat

Sulfatreat is a fixed bed process designed to remove hydrogen sulphide and mercaptans from natural gas.

Equipment is simple and consists of an upstream separator and the reactor itself. The catalyst is disposed when exhausted and as feed concentration is increased up to 2800 ppm, catalyst life is reduced. Almost 800 lbs of H₂S can be adsorbed before change-out is required.

Over 100 plants are in operation, and Shell have replaced two iron oxide facilities in Michigan, USA.

The catalyst is non-toxic, non-phosphoric and environmentally safe. PH control is not required and the process generates no toxic or corrosive gas.

There is little chance of methane hydrate formation due to low pressure drop across the bed. Natural Gas Sweeteners Association licenses the process.

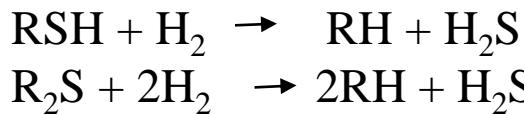
Batch Solvents/Beds

Chemsweet

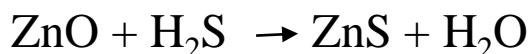
Chemsweet is a process that removes hydrogen sulphide from natural gas using a slurry suspension of fine zinc oxide particles in zinc acetate solution.

Mercaptans and carbonyl sulphide are also removed by reaction with the slurry to form zinc sulphide and zinc mercaptide. A dispersant is added to the slurry to keep solid particles in suspension, spent slurry passes to a degassing tower and is dumped.

Desulfurization of organic sulphur compounds is typically accomplished, by first, hydrogenating these compounds to form hydrogen sulphide:



Then, the hydrogen sulphide reacts with the zinc oxide adsorbent to form zinc sulphide, effectively trapping the sulphur:



Batch Solvents

Puraspec

The puraspec process licensed by Synetix, can be used for removal of ppm concentrations of H₂S and COS from hydrocarbon gases and liquids to meet pipeline or petrochemical specifications as required.

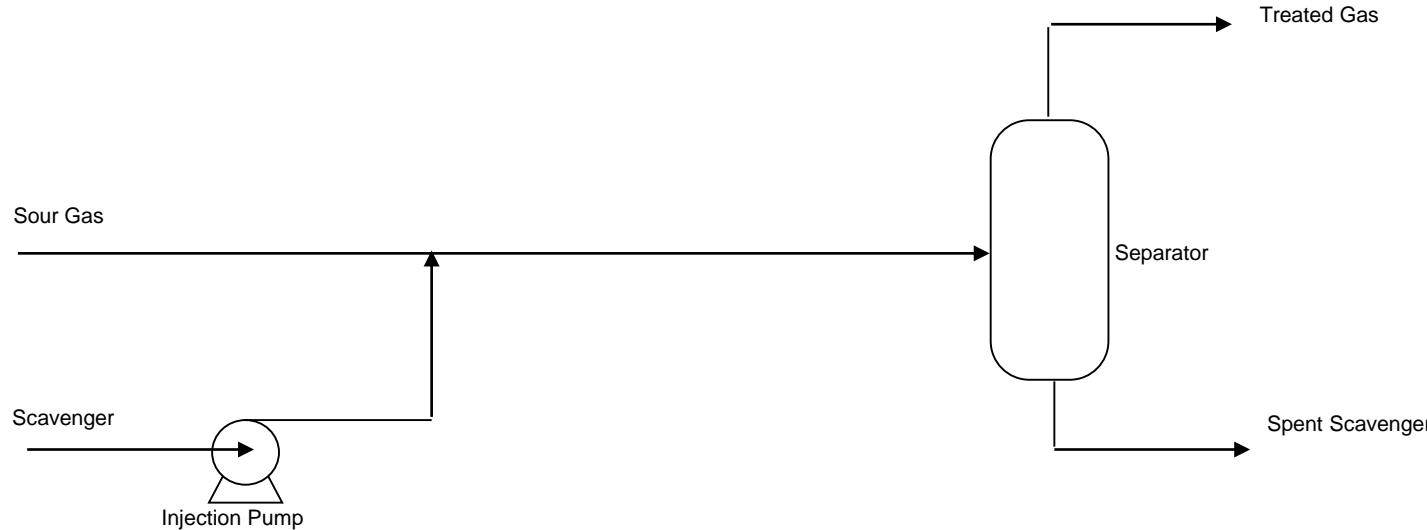
Fixed beds of chemical absorbents provide effectively total irreversible selective removal of impurities from wet or dry hydrocarbons without feedstock losses.

The operating conditions can range from: temperatures of -7°C to 204 °C; and pressures of atmospheric to 138 bar.

Typically the process uses two separate reactors so that continuous production can be maintained even when one bed is being recharged.

Can be used to remove mercury.

H₂S Scavengers

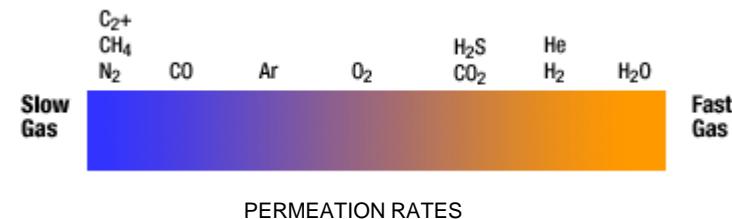
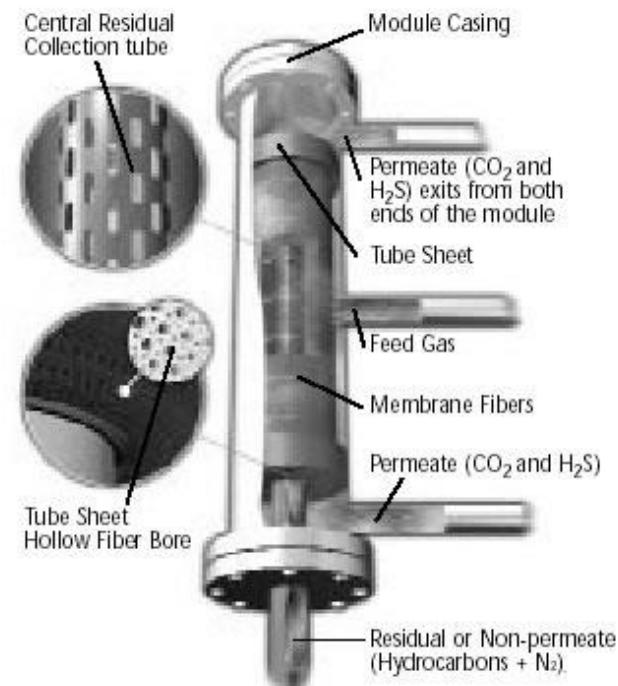


A direct injection H₂S scavenging system for natural gas is very simple, consisting simply of an injection pump, injection point, a length of pipe for scavenger-gas contact, and a means of separating the spent scavenger (normally provided by existing equipment).

The method of introducing the scavenger into the gas line may either be piping tee, injection quill, or atomising nozzle. In-line mixers have also been used downstream of the injection point in order to ensure proper gas – scavenger contact. Most scavenging agents are designed to treat gas over a wide range of temperature, pressure and flow conditions. However, there are a number of issues which limit the suitable injection locations:

Membrane Processes

Membrane separation is based on the selective diffusion of gases through a permeable membrane. Gas mixtures are separated on the basis of their different solubility and diffusivity through the membrane barrier when under an imposed partial pressure gradient, with the feed gas at high pressure and the permeate gas at low pressure. The components of the natural gas diffuse at different rates, i.e. H_2O vapour, CO_2 and H_2S have high permeation rates, methane has a medium rate and heavier hydrocarbons like butane have a much lower permeation rate.



Mercury Removal

H_2S (traces) removal on metal oxide

- Acid gas content <1%
- Chemical reaction



- <0.1ppmv H_2S in product gas
- Non-regenerative



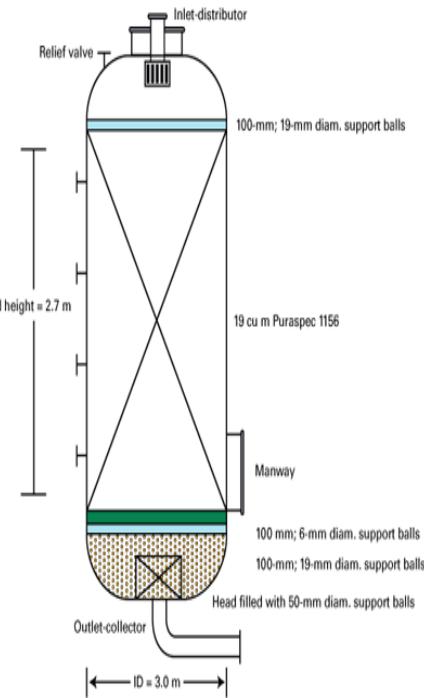
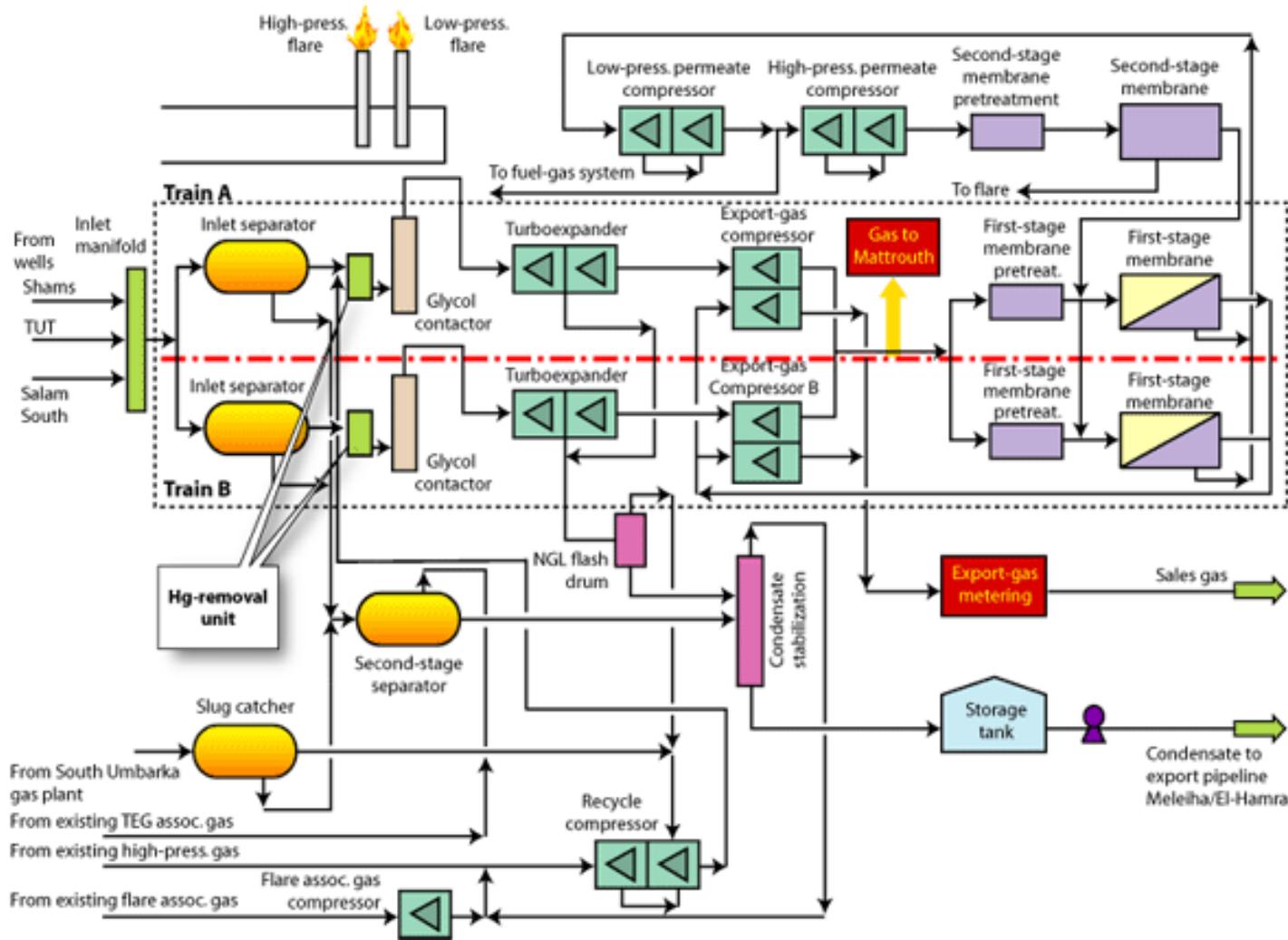
Upstream precautions:

- No liquids (free water / condensate)
- Minimise glycols, amines and particles

Downstream effects:

- Water
- Dust particles

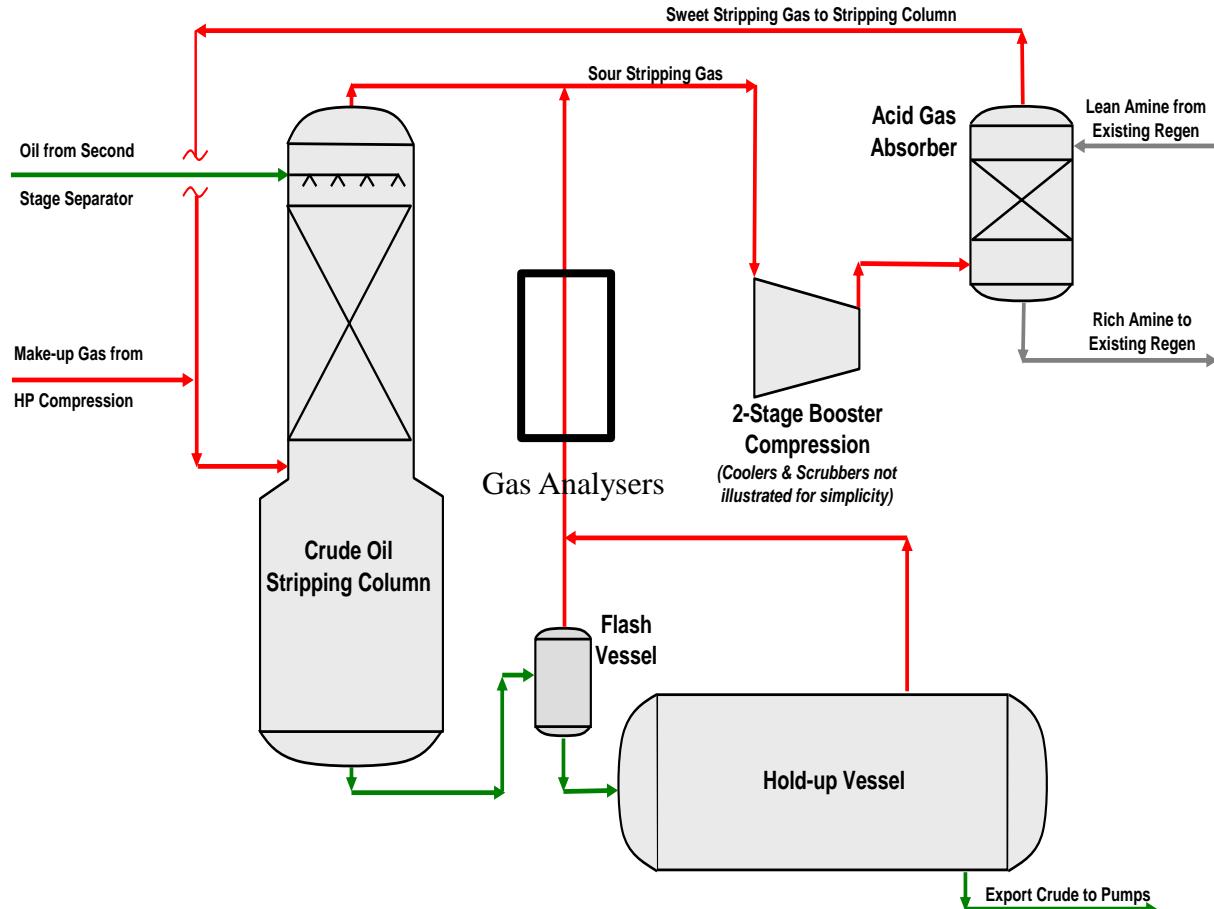
Salam Gas Plant - Egypt



H₂S Oil Stripper Desorption Column Typical Arrangement

Oil Flow Path:

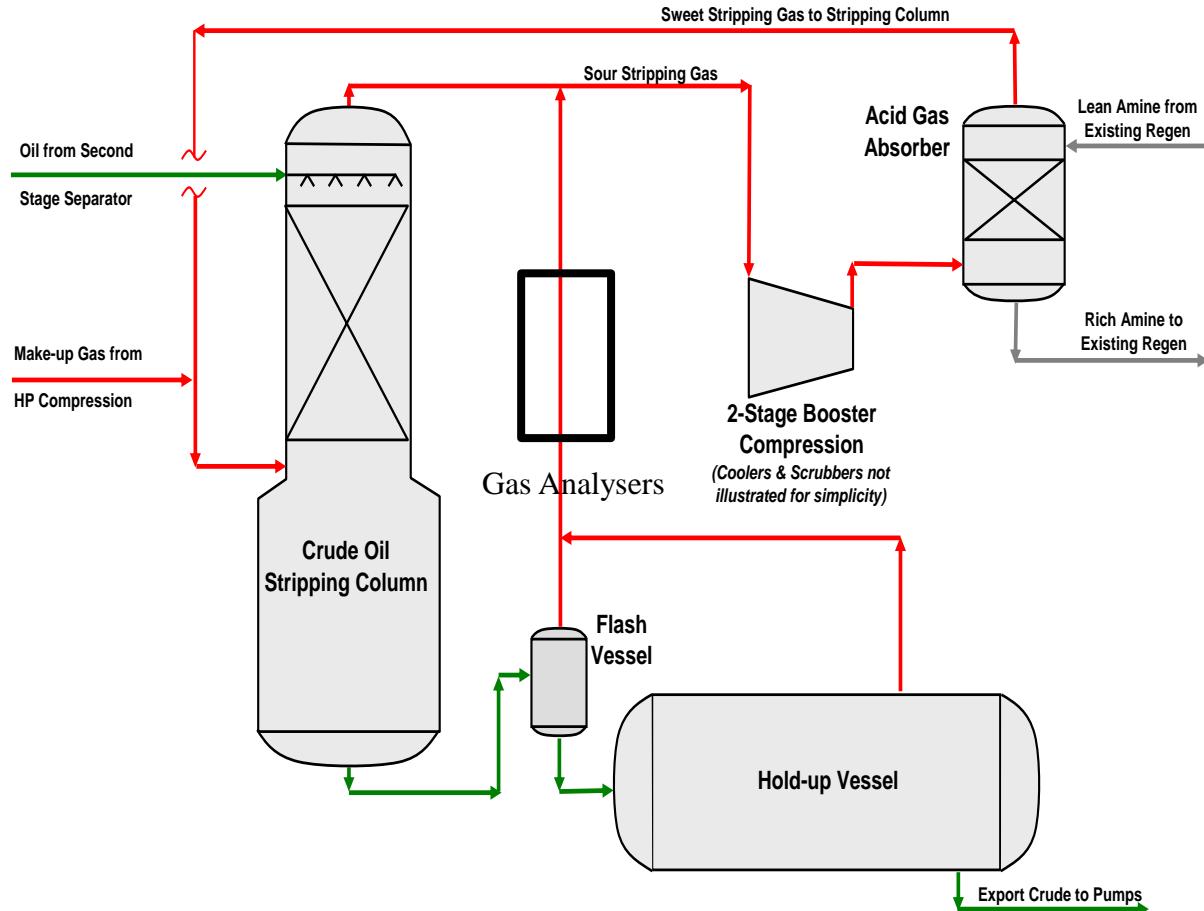
1. The Oil is fed into the desorption column via a nozzle array.
2. Oil then progresses through the plate / packing section and is contacted counter-currently with stripping gas to remove H₂S.
3. Treated sweet oil ($\leq 11\text{ppm}$) is collected at the bottom of the desorption column.
4. Oil passes through the flash vessel to provide the gas analysers with an H₂S sample.
5. The oil is then stored in the hold-up vessel for approximately 3 minutes whilst the gas analysers recorded the results of the sample.
6. Oil stored in the hold up vessel is then exported to the Oil Pipeline.



Desorption Column Typical Arrangement

Gas Flow Path:

1. Production gas (considered to be sweet) is fed into the system.
2. Gas passes up through the desorption column and is contacted counter-currently with the crude oil along the active tray area.
3. Sour stripping gas exits the desorption column and passes through the 2-stage booster compression.
4. Sour stripping gas is sent to the acid gas absorber to remove H₂S using Amine.
5. Note - Designed outlet specification is 10ppm H₂S.
6. The sweet stripping gas is recycled back into the desorption column and if required secondary supply of stripping gas is available from existing gas production.



Sulphur Recovery

General

Sulphur is present in natural gas principally as hydrogen sulphide (H_2S). The H_2S present is removed from natural gas by means of one of the acid gas treatment processes described earlier. The resulting H_2S -containing acid gas can be converted to elemental sulphur by means of the Claus process.

Claus Process

The Claus process is split into two parts.

1. A thermal or combustion reaction section in which most of the overall heat of reaction (from burning $\frac{1}{3}$ of the H_2S and essentially 100% of any hydrocarbons and other combustibles in the feed) is released and removed

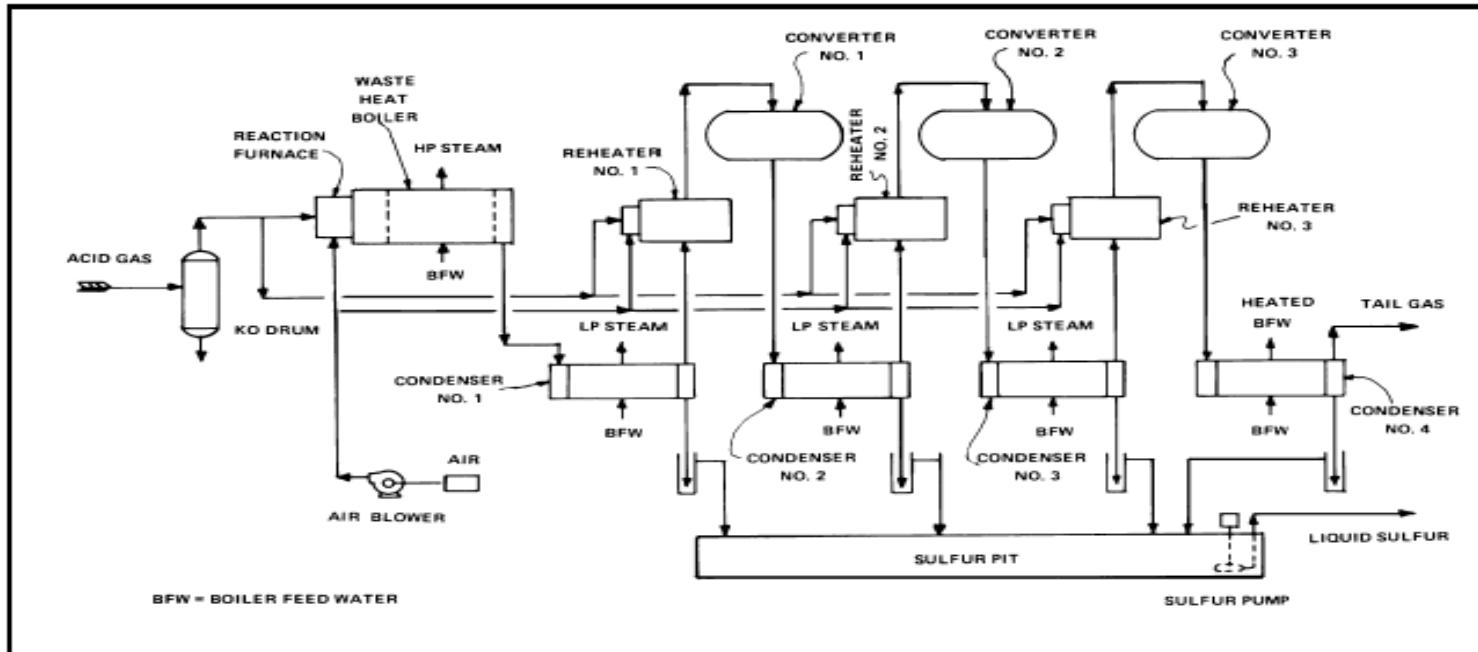


2. A moderately exothermic catalytic reaction section in which sulphur dioxide (SO_2) formed in the combustion section reacts with unburned H_2S to form elemental sulphur.



Claus Process

Three-Stage Sulphur Plant



In this method, the re-heaters are fired by acid gas. As an alternative, treated gas and air could be used, lowering sulphur emissions in the process.

Sulphur Recovery

The Kaybob Sulphur Recovery Plant, Alberta, Canada



Key Learnings

- Gas specifications – water, cricondenbar, calorific value, wobbe index
- Water content of gas
- Hydrate formation
- Principles of glycol dehydration and construction of operating and equilibrium lines
- Principles of molecular sieve
- Refrigeration, Joule-Thomson and turbo-expansion processes
- Use of Mollier chart to estimate refrigeration system components
- Principles of amine absorption for CO₂ and H₂S removal.