University of Calgary

Natural Gas Processing Principles and Technology - Part II

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Purpose of this Document

1. To study the physical, chemical and engineering principles used in the handling of natural gas, substitute natural gas, their associated liquids, and the other compounds obtained from them.
2. To apply these principles to actual design and operating problems.

Preface

The gas processing industry in the past 25 years has developed extensively. Many different qualities of gas have been discovered and are being processed. Many different processing schemes are used to accomplish the same ends. This book tries to outline the major ways that this is being done, and to make the reader aware of the main areas of technology that are considered in gas processing and handling.

This book is intended for practising engineers of all disciplines as a guide for them in the gas processing industry as gas engineers come from all branches of engineering.

At the end of most chapters, typical problems that might be encountered in industry are given. As these are practical problems, more than the principles of the particular chapter have to be used to solve the problems.

The Gas Processors Suppliers Association Engineering Data Book1 has been referred to extensively for data and other material as it is the basic reference book for the industry.

This book has been developed from the lecture notes developed by the author for a course given by him at the University of Calgary in the Department of Chemical Engineering over the past 24 years, and at other centres in both North America and Europe. During that period, several people have lectured in various sections of the course and have helped in the preparation of the text. These are, with the address when they participated in the course or their present address are as follows:

Dr. N. Anderson, Kilborn Engineering, Toronto, Ontario

Dr. K. Aziz, Stanford University, Palo Alto, California

Mr. E. Berlie, Western Research & Development, Calgary, Alberta

Dr. A. Chakma, University of Calgary, Calgary, Alberta

Dr. P.R. Bishnoi, University of Calgary, Calgary, Alberta

Dr. K. Goddard, Chevron Standard, Calgary, Alberta

Dr. G.A. Gregory, Neotechnology Consultants Ltd., Calgary, Alberta

Mr. G.E. Handwerk, Consultant, Denver, Colorado

Dr. R.A. Heidemann, University of Calgary, Calgary, Alberta

Mr. G. Kinsep, SKM Consulting, Calgary, Alberta

[1] Published by the Gas Processors Suppliers Association and the Gas Processors Association, Tulsa, Oklahoma. Latest edition is the 10th edition which is published in English Units only. Page No.'s in the text refer to the Metric edition or the 9th English edition unless otherwise noted.

Mr. M.A. Kyllo, Ralph M. Parsons Company Ltd., London, England

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Mr. E. Wichert, Petro Canada, Calgary, Alberta

Also, as the notes are based on the author's more than 30 years in the gas processing industry, he wishes to acknowledge the many persons not already acknowledged who contributed data and suggestions incorporate in the book.

The more recent revisions have been made to bring the notes up—to—date and to show both metric units and English units whenever appropriate. The problem of using metric units is that the source of much of the data is the 9th edition of GPSA Metric Data book, which is based on 1972 data. There is now available a 10th edition of this book which was published in 1987 that gives English units only. This is a much more complete reference book and as such will be used extensively in class.

10.0 Compression and Refrigeration

- A review of the theory of compression and refrigeration operations.
- The sizing equipment for compression and refrigeration operations.

10.1 Compression

10.1.1 Types

There are several types of compressors. Table 10.1 gives the main types and their general application; ie. reciprocating, rotary centrifugal or, axial flow types. The types and application are shown on Figure 10.2.

10.1.2 Determining Discharge Pressure

This is done as outlined in Chapter 6, Fluid Flow. A diagram of the system is drawn and the total pressure at the discharge of compressor is worked out by calculating the pressure drop in the discharge system of the compressor. The following example shows how this is done.

Example Problem 10.1

Determine the discharge pressure required and the compression ratio for the compressor shown in Figure 10.1.

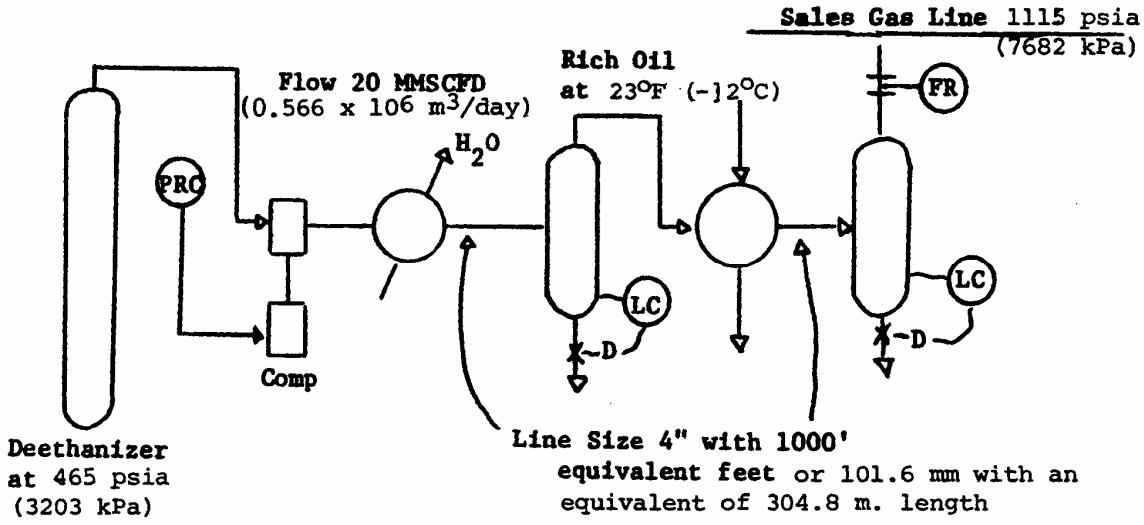


Figure 10.1: Compressor Discharge System

Table 10.1: Compressor Types and Ranges

Type	Maximum Pressure Range PSIG	Approx. Maximum Compression Ratio/Stage	Maximum Compression Ratio	Application	Example In Gas Processing Industry
Reciprocating	30,000 - 50,000	10	As required	All flows & pressures	Reinjection pipeline. Stations

					gathering systems. Small refrigeration loads.
Centrifugal	3,000 - 5,000	3	6	High Flow	Large refrigeration units. Pipe-Trans. Comp.
Rotary Displacement	130	4	4	High Flow Low Press.	Stock Recovery Units
Axial	3,000 - 5,000	1.5	5.65	High Flow Low Press. Differential	Sulphur Plant Blowers. Very large HP in LNG plants
Rotary Screw	800	10	As required	Constant flow medium quantity	Propane Refrigeration Instrument Air

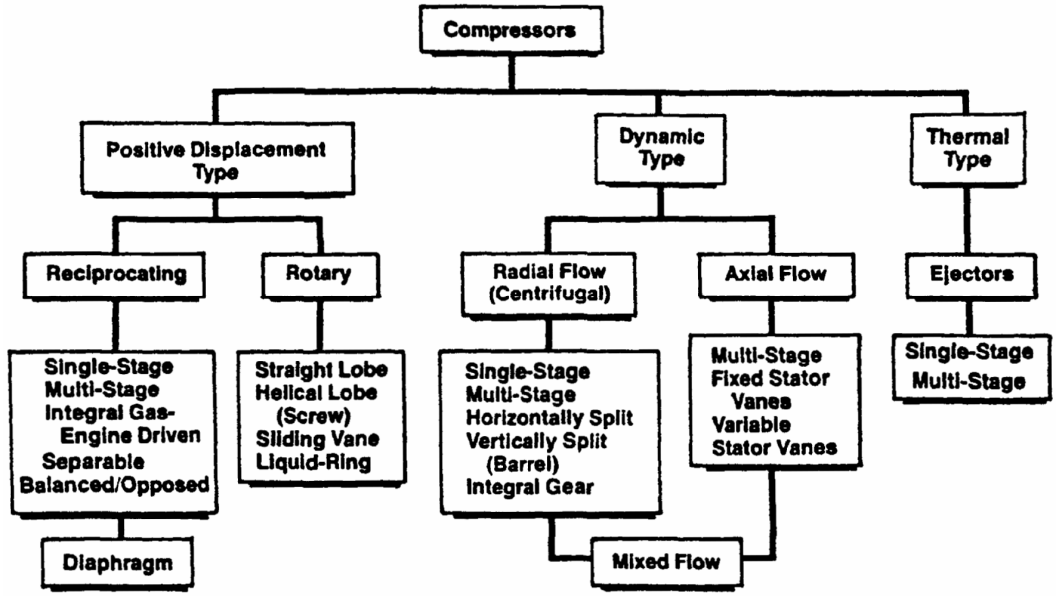


Figure 10.2: Types of Compressors

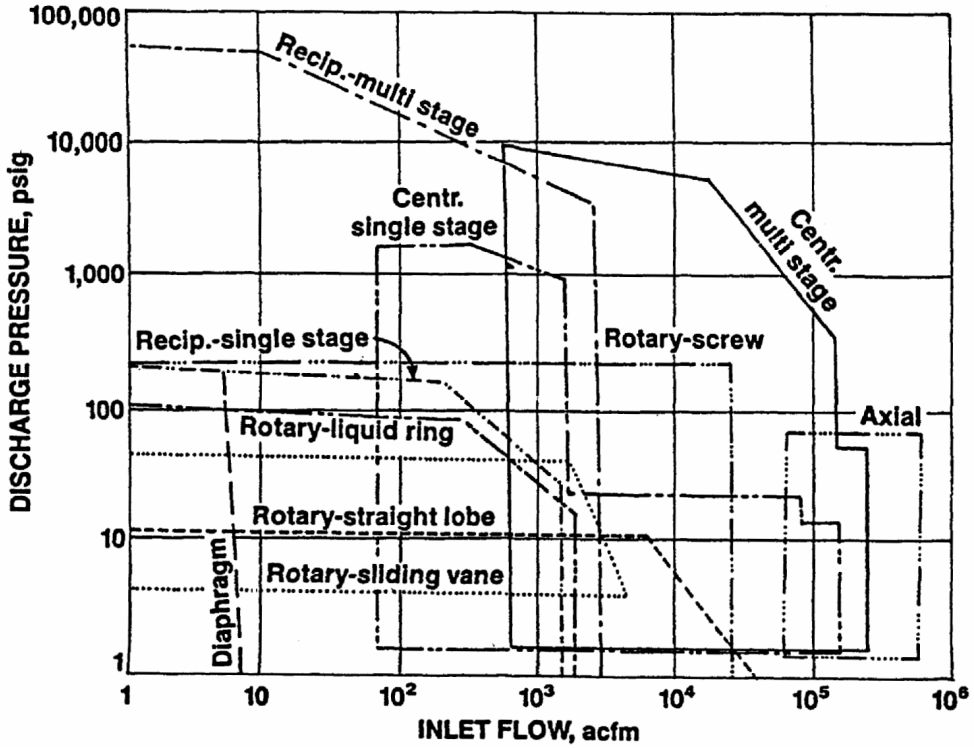


Figure 10.3: Compressor Coverage Chart

	English Units	Metric Units
	PSIA	kPa
Suction Pressure	465	3203
Final Pressure	1115	7682
Pressure drop through exchangers (10 psi/exchanger)	20	128
Flow Recorder	3	21
Piping Losses	30	207
Safety Factor 32 of Difference between Suction S Discharge Pressure .03 (700)	21	145
Total Discharge Pressure	1189	8193
Compression Ratio	$\frac{1189}{465} = 2.56$	$\frac{8193}{3203} = 2.56$

10.1.3 Work of Compression

The work required to compress a gas from Condition 1 to Condition 2 can be found from the energy equation which was given in Fluid Flow.

Equation 10-1

$$U_2 + \frac{v_2^2}{2g_C} + Z_2 + P_2V_2 = U_1 + \frac{v_1^2}{2g_C} + Z_1 + P_1V_1 + Q + W$$

or

Equation 10-2

$$\Delta U + \frac{\Delta\left(v_2^2 - v_1^2\right)}{2g_C} + \Delta Z + \Delta PV = Q + W$$

where:

U = the internal energy.

v = the velocity.

Z = the height or potential.

P = the pressure.

W = the work.

Q = the heat added to the system.

V = the volume.

For normal calculations the changes in potential, kinetic energy, etc., are neglected. We then get two cases. These are both based on integrating the change of P and V between points 1 and 2.

Equation 10-3

$$W = \int_P^{P_2} V dP$$

Isothermal Compression (at constant temperature)

The ideal gas law equation applies. The work required in terms of HP:

Equation 10-4

$$HP = 0.004363 P_1 V_1 \ln\left(\frac{P_2}{P_1}\right)$$

where:

P_1 = the intake condition.

P_2 = the final conditions.

V_1 = the intake volume in CF/min (actual CFM).

P = the inlet pressure in PSIA.

In terms of thermodynamic expressions, for isothermal compression we have the theoretical work expressed as:

Equation 10-5

$$Work = \Delta H - T\Delta S$$

or for 1 MMSCFD of gas

Equation 10-6

$$HP = 0.0432(\Delta H - T\Delta S)$$

where:

ΔH = the enthalpy change in BTU's/lb mole.

ΔS = the entropy change in BTU's/lb mole, ºR.

Adiabatic Compression

No outside heat enters or leaves. In this case gas obeys the expressions PVK = constant. For this the theoretical HP can be expressed by:

Equation 10-7

$$HP = 0.004363 P_1 V_1 \left[\left(\frac{K}{K-1} \right) \left(\frac{P_2}{P_1} \right)^{K-1/K} - 1 \right]$$

where:

V_1 = cu. ft/min at intake conditions (actual CFM).

K = the ratio of specific heats $\frac{C_P}{C_V}$.

P_1 = the suction pressure psia.

P_2 = the discharge pressure psia.

Most compressors operate close to the adiabatic case. For adiabatic compression the work is expressed by:

Equation 10-8

$$Work = \Delta H$$

where:

ΔS = 0, ie. constant entropy expression.

Therefore for 1/MMSCFD:

Equation 10-9

$$HP = 0.0432\Delta H$$

where:

ΔH is in BTU/lb mole.

Calculating the HP Required

As a quick approximation, use for reciprocating compressors.

$$HP = 22 \left(\frac{Compression\ Ratio}{Stage} \right) (No.\ of\ \ Stages)(F)\, MMCFD$$

or in metric units

$$W = 14\left(\frac{Compression\ Ratio}{Stage}\right)(No.\ of\ Stages)(F)\,m^3/hr$$

where:

F = 1.00 for 1 stage.

1.08 for 2 stages.

1.10 for 3 stages.

W = watts.

See pages 13.4 (4-2) and Fig. 13.9 (4.4) of the GPSA Data Book for further details.

Note: Figure no's and pages in booklets are from Metric Data Book. Others are from 1987 English Data Book.

For a more accurate means of calculating the HP charts are given in the GPSA Data Book plotting BHP vs ratio of specific heat (K) and the compressor ratio. The mechanical efficiency is allowed for in the charts. However corrections have to be made for specific gravity, which is given in Fig. 13-15 (4-12) in the GPSA Data Book and correction factor for low intake pressure is given by Fig. 13-12 (4-11). Also other corrections have to be made as shown below.

The horsepower for compression is found from:

Equation 10-10

$$BHP = BHP\Big/ MMSCFD\left(\frac{P_L}{14.4}\right)\left(\frac{T_S}{T_L}\right)(Z_{ave}) \qquad (Flow\ in\ MMSCFD)$$

or

$$W = \frac{W}{m^3/hr}\left(\frac{P_L}{101.325}\right)\left(\frac{T_S}{T_L}\right)(Z_{ave}) \qquad (Flow\ in\ m^3/hr)$$

where:

BHP/MMSCFD or $\frac{W}{m^3/hr}$ is found from charts (4-7) through (4-10) and corrected if necessary by charts 13-12 (4-11) and 13-15 (4-12).

P_L = Pressure base psia or kPa.

T_S = Intake temperature ºR or K.

T_L = Temperature base ºR or K.

MMSCFD = Flow at standard conditions of the base given by P_L + T_L.

(m^3 /hr) = Flow at standard conditions of the base given by P_L + T_L.

Z_{ave} = Average compressibility factor in the cylinder $\frac{Z_{suction} + Z_{discharge}}{2}$.

Please note this data is for slow speed units. For high speed units make the corrections shown on Page 4.13 of the GPSA Data Book.

10.1.4 Compression Ratio

$$\frac{Discharge\ Absolute\ Pressure}{Suction\ Absolute\ Pressure}$$

Normally do not use a compression ratio of greater than 6 for single stage compression. Generally it is around 3.

10.1.5 Ratio of Specific Heats

In the equation for adiabatic compression the term K is used. It is the ratio of the specific heat at constant pressure (Cp) and specific heat at constant volume (CV). It can be calculated if the value of the specific heat at constant pressure is known from the equation

Equation 10-11

$$MC_P - MC_V = 1.99$$

where:

M = molecular weight.

Therefore

Equation 10-12

$$K = \frac{MC_P}{MC_P - 1.99}$$

in metric units

$$K = \frac{MC_P}{MC_P - 8.3145}$$

The GPSA Data Book, Fig. 4-1, gives the value of MC for most compounds P expected in natural gas at various temperatures so that a K for a mixture can be easily calculated. Some common values of K are:

for Air 1.4

for Natural Gas 1.26

For a quick calculation of "K" for hydrocarbon gases, use Fig. 4-3 of the GPSA Data Book.

The value of K is one of the most important values in determining the compression power required. The following example problem illustrates the calculations required.

Example Problem 10.2

Calculate the HP required to compress 80 MMSCFD of gas of the composition given below from 400 psia at 100ºF to 1100 psia.

First we have to calculate the ratio of specific heat $"K" = \frac{C_p}{C_V}$.

This is found from the molal heat capacity for the mixture.

Component	Mole %	Mole* Heat Cap	Partial Heat Cap	MW	pMW
H_2S	2.4	8.27	0.20	34	0.82
CO_2	4.4	9.28	0.41	44	1.93
C_1	85.4	8.95	77.64	16	13.66
C_2	4.5	13.77	0.62	31	1.35
C_3	1.6	19.52	0.31	44	0.70
iC_4	0.4	25.77	0.10	58	0.23
nC_4	0.7	25.81	0.18	58	0.40
iC_5	0.6	31.66	0.19	72	0.43
			9.65		19.52

* At 150ºF average temperature approximately the average temperature in the compressor. Calculate the discharge temperature using the graph, Fig. 4-6 of the GPSA Data Book and find the arithmetic average between suction and discharge conditions.

You Can also use chart in GPSA Data Book for this calculation.

To find the compression HP, use the graph in the GPSA Data Book and find that the HP per MMCF at a compression ratio of

$$\frac{1100}{400} = 2.75$$ and a "K" of 1.26 is 59 or in metric units 38.8 W/m3/hr.

$$K = \frac{9.65}{9.65 - 1.99} = 1.26$$

Corrections for intake pressure and for specific gravity are both negligible and multiplying factors are 1.0.

$$\frac{T_S}{T_L} = \frac{560}{520} = 1.076$$

Now find Z_{ave}

$$Z_{400} PSIA = .93$$
$$\alpha 100°F$$
$$Z_{1100 PSIA \alpha 226°F} = .94$$
$$Z_{ave} = .935$$
$$BHP = 59(1.076)(.935)80$$
$$= 4748$$

By the Cooper Bessemer slide rule the HP required works out to 56/MMSCF or 4500 HP, which is a good check.

10.1.6 Reciprocating Compressor Capacity

The capacity of a reciprocating compressor is a function of the suction pressure, the piston displacement and the strokes per minute. Figure 10.4 below illustrates a problem that occurs.

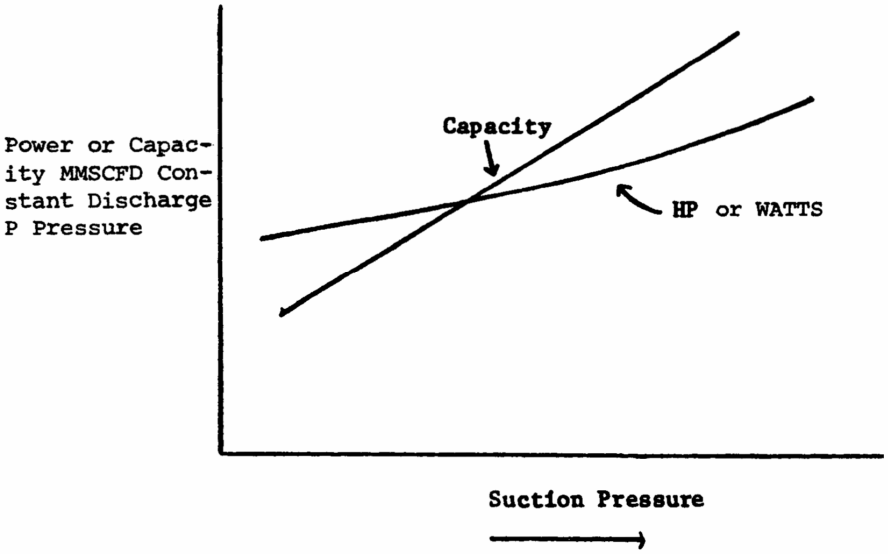


Figure 10.4: Capacity and Power of Reciprocating Compressors

As the suction pressure is dropped the capacity of the machine as represented by the Stroke-Piston displacement and RPM decreases faster than the HP required for the compression ratio, and the volume that can be pumped. The machine does not have enough HP and the compressor stalls or volumetric changes to reduce the capacity have to be made. (i.e., clearance pockets).

The flow as mentioned is a function of Piston Displacement. For a single acting compressor:

Equation 10-13

$$P_D = \frac{A_{HE}(S)(RPM)}{1728}$$

or

$$P_D = \frac{A_{HE}(S)(RPM)(60)}{4\ x\ 10^9}$$

and for a double acting compressor:

Equation 10-14

$$P_{DA} = \frac{(S)(RPM)(2)}{1728}\left[A_{HE} - \frac{A_R}{2}\right] = 4.55\ x\ 10^{-4}(S)(RPM)(2D^2 - d^2)$$

or

$$P_{DA} = \frac{(S)(RPM)\,2\left(A_{HE} - A_R/2\right)}{4\ x\ 10^9}$$

where

P_D = Piston Displacement in CFM (single acting) or m^3/hr.

P_{DA} = Piston Displacement in CFM (double acting) or m^3/hr.

S = Stroke in inches or mm.

RPM = Revolution per minute.

A_{HE} = Area head end in inches^2 or mm^2.

A_R = Area rod in inches^2 or mm^2.

ACE = Area of crank end ACE = A_{HE} - A_R in inches^2 or mm^2.

D = cylinder diameter in inches^2 or mm^2.

d = piston rod diameter in inches^2 or mm^2.

Now the capacity of a compressor is found in SCFD per day from

Equation 10-15

$$Capacity = P_D(E_V)\left(\frac{P_1}{14.4}\right)(1440)\left(\frac{520}{T_S}\right)\left(\frac{1}{Z}\right)$$

or simplifying for ordinary temperatures and low pressures in English Units.

Equation 10-16

$$Capacity = P_D E_V P_1(100)$$

in Metric Units

$$m^3/hr = \frac{P_D E_V P_1}{(100)(101.325)Z_{ave}}$$

where

P_1 = the suction pressure.

E_V = the volumetric efficiency which is found from

Equation 10-17

$$E_V = 96 - R - \%Cl\left[\left(\frac{Z_S}{Z_d}\right)(R)^{1/K} - 1\right]$$

where

$$\%Cl = Clearance\ Volume = \frac{V_C\ (clearance\ \ volume)}{P_D\ (displacement)}$$

$$R = compression\ ratio\ \frac{P_2}{P_1}$$

10.1.7 Rod Loading

The net or differential force in pounds exerted upon a piston rod due to gas pressure differences between head end of a piston and the rod is called the rod load. Generally it is calculated as follows:

Equation 10-18

$$Rod\ Load\ in\ compression = A_{HE}P_2 - A_{CE}P_1$$

The compressor manufacturer states the allowable rod loading for the machine and the calculated value should be well below this.

10.1.8 Specifying Reciprocating Compressors

There are several items that should be specified or obtained from the manufacturer when receiving a bid for the compressor. These include:

Table 10.2: Specifications for Reciprocating Compressors

	Specified By
The capacity required	Operator
The gas being compressed	Operator
The suction pressure	Operator
The suction temperature	Operator
The discharge pressure	Operator
The cylinder material	Operator or Manufacturer
The liner material	Operator or Manufacturer
Unloading Facilities or the method of flow control	Operator or Manufacturer
Lubrication of piston required	Operator
Rod Loading allowed	Manufacturer
Allowable vibration in piping	Operator
Driver type and speed	Operator
No. of stages	Manufacturer
Discharge temperature	Manufacturer
HP required for the service	Manufacturer
Intercoolers required and supplied	Manufacturer
Type of machine	Operator or Manufacturer
Allowable piston speed	Operator
Valve type and material	Manufacturer
Coupling to driver	Manufacturer
Area of installation and elevation	Operator
Cooling requirements	Manufacturer

Considerably more data has to be supplied if the compressor and driver are a single package. Also, it is generally desirable to specify a standard specification such as API 618 which covers reciprocating compressors.

10.1.9 Trouble Shooting Reciprocating Compressors

There are many operating problems that can occur with reciprocating compressors. Table 10.3 from the latest GPSA Data Book (English units version) outlines many of the problems and their causes.

Table 10.3: Probable Causes of Reciprocating Compressor Double

Trouble	Probable Causes
Compressor will not start	1. Power supply failure.
	2. Switchgear or starting panel.
	3. Low oil pressure shut down switch.
	4. Control panel.
Motor will not Synchronize	1. Low voltage.
	2. Excessive starting toque.
	3. Incorrect power factor.
	4. Excitation voltage failure.
Low oil pressure	1. Oil pump failure.
	2. Oil foaming from counter weights striking oil surface.
	3. Cold oil.
	4. Dirty oil filter.
	5. Interior frame oil leaks.
	6. Excessive leakage at bearing shim tabs and/or bearings.
	7. Improper low oil pressure switch setting.
	8. Low gear oil pump by-pass/relief valve setting.
	9. Defective pressure gauge.
	10. Plugged oil sump strainer.
	11. Defective oil relief valve.
Noise in cylinder	1. Loose piston.
	2. Piston hitting outer head or frame end of cylinder.
	3. Loose crosshead lock nut.
	4. Broken or leaking valve(s).
	5. Worn or broken piston rings or expanders.
	6. Valve improperly seated/damaged seat gasket.
	7. Free air unloader plunger chattering.
Excessive packing Leakage	1. Worn packing rings.
	2. Improper lube oil and/or insufficientlube rate (blue rings).
	3. Dirt in packing.
	4. Excessive rate of pressure increase.
	5. Packing rings assembled incorrectly.
	6. Improper ring side or end gap clearance.
	7. Plugged packing vent system.
	8. Scored piston rod.

Trouble	Probable Causes
Packing over heating	1. Lubrication failure.
	2. Improper lube oil and/or insufficient lube rate.
	3. Insufficient cooling.
Excessive carbon on valves	1. Excessive lube oil.
	2. Improper lube oil (too light, high carbon residue).
	3. Oil carryover from inlet system or previous stage.
	4. Broken or leaking valves causing high temperature.
	5. Excessive temperature due to high pressure ratio across cylinders.
Relief valve popping	1. Faulty relief valve.
	2. Leaking suction valves or rings on next higher stage.
	3. Obstruction (foreign material, rags), blind or valve closed in discharge line.
High discharge temperature	1. Excessive ratio on cylinder due to leaking inlet valves or rings on next higher stage.
	2. Fouled intercooler/piping.
	3. Leaking discharge valves or piston rings.
	4. High inlet temperature.
	5. Fouled water jackets on cylinder.
	6. Improper lube oil and or lube rate.
Frame knocks	1. Loose crosshead pin, pin caps or crosshead shoes.
	2. Loose/worn main, crankpin or crosshead bearings.
	3. Low oil pressure.
	4. Cold oil.
	5. Incorrect oil.
	6. Knock is actually from cylinder end.
Crankshaft oil seal leaks	1. Faulty seal installation.
	2. Clogged drain hole.
Piston rod oil scraper leaks	1. Worn scraper rings.
	2. Scrapers incorrectly assembled.
	3. Worn/scored rod.
	4. Improper fit of rings to rod/side clearance.

Courtesy of Ingersoll-Rand Co.

10.1.10 Centrifugal Compressors - HP or W Calculations

Centrifugal compressors are used for high flow and relatively low compression ratios. Figure 10.6 shows the different characteristics of centrifugal axial and reciprocating compressors. In considering compression for these machines, the compression law follows an equation similar to the reciprocating machines. To calculate the BHP the approximate equation can be used:

Equation 10-19

$$BHP/MMSCFD = 240.5\left[(R)^{\frac{K-1}{K}} - 1\right]$$

where

R = the compression ratio per unit or case.

This is assuming an efficiency of 73%. A more general equation is:

Equation 10-20

$$BHP = \frac{1545(Z_a)T_1W}{33000\,E(MW)}\left(\frac{K}{K-1}\right)\left[(R)^{\frac{K-1}{K}} - 1\right]$$

where:

Za = average compressibility factor.

T_1 = suction temperature ºR.

W = gas flow rate lb/min.

E = overall efficiency.

MW = molecular weight.

In metric units the power in KW is given by:

$$Power = \frac{1000W\left(\frac{8.314}{M}\right)\left(\frac{K}{K-1}\right)(Z_{ave}T_1)\left[\left(\frac{P_2}{P_1}\right)^{\frac{K-1}{K}} - 1\right]}{3600000(E)}$$

where:

W = flow in kg/hr.

M = mole wt kg.

K = ratio of specific heats.

Z_{ave} = average compressibility factor.

P_2 = final pressure kpa.

P_1 = initial pressure kpa.

T_1 = initial temperature ºk.

E = efficiency.

10.1.11 Centrifugal Compressors - General Operating Characteristics

The centrifugal (and the axial compressor) are generally like a centrifugal pump. They develop pressure from rotating blades or impellers and they have a performance curve as shown in the figure below.

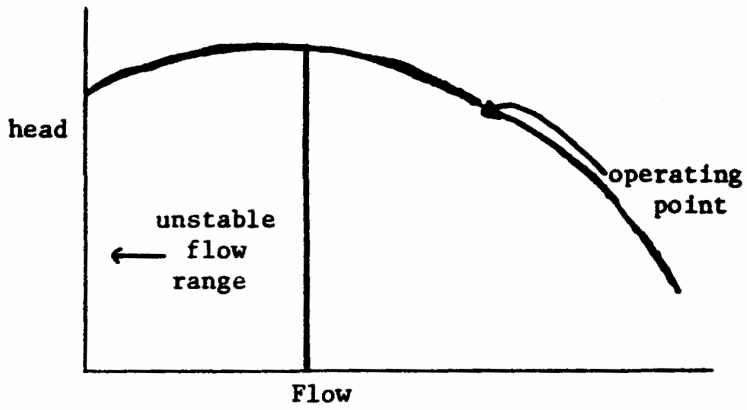


Figure 10.5: Head Capacity Curve for Centrifugal Compressors

The system and the compressor characteristics have to meet. However the system can change by changing pressures and of course the compressor curve can be changed in many ways such as changing the speed, etc. The head required by the system is calculated from the following equation:

Equation 10-21

$$H_{adiabatic} = ZRT\left(\frac{K}{K-1}\right)\left[\left(\frac{P_2}{P_1}\right)^{\frac{K-1}{K}} - 1\right]$$

The compressor head developed is found from the equation

Equation 10-22

$$H_{adiabatic} = \psi\frac{u^2}{g}$$

where

ψ = head coefficient and is a function of blade angle, efficiency, etc. It is normally 0.5 for industrial compressors and goes to 0.63 with radial blading.

u = tip speed ft/sec.

g = gravity.

Each stage of a compressor normally runs between 8000 to 12,000 feet of adiabatic head.

Figure 10.7 is a typical performance map which shows the basic shape of performance curves for a variable speed centrifugal compressor. The curves are affected by many variables, such as desired compression ratio, type of gas, number of wheels, sizing of compressor, etc.

With variable speed, the centrifugal compressor can deliver constant capacity at variable pressure, variable capacity at constant pressure, or a combination variable capacity and variable pressure.

Basically the performance of the centrifugal compressor, at speeds other than design, is such that the capacity will vary directly as the speed, the head developed as the square of the speed, and the required horsepower as the cube of the speed. As the speed deviates from the design speed, the error of these rules, known as the affinity laws, or fan laws , increases.

Fan Laws

Equation 10-23

$$Q \propto N; \; i.e. \; \frac{Q_{110}}{N_{110}} = \frac{Q_{100}}{N_{100}} = \frac{Q_{90}}{N_{90}}$$

$$H \propto N^2; \; i.e. \; \frac{H_{110}}{(N_{110})^2} = \frac{H_{100}}{(N_{100})^2} = \frac{H_{90}}{(N_{90})^2}$$

$$Bhp \propto N^3; \; i.e. \; \frac{Bhp_{110}}{(N_{110})^3} = \frac{Bhp_{100}}{(N_{100})^3} = \frac{Bhp_{90}}{(N_{90})^3}$$

By varying speed, the centrifugal compressor will meet any load and pressure condition demanded by the process system within the operating limits of the compressor and the driver. It normally accomplishes this as efficiently as possible, since only the head required by the process is developed by the compressor. This compares to the essentially constant head developed by the constant speed compressor.

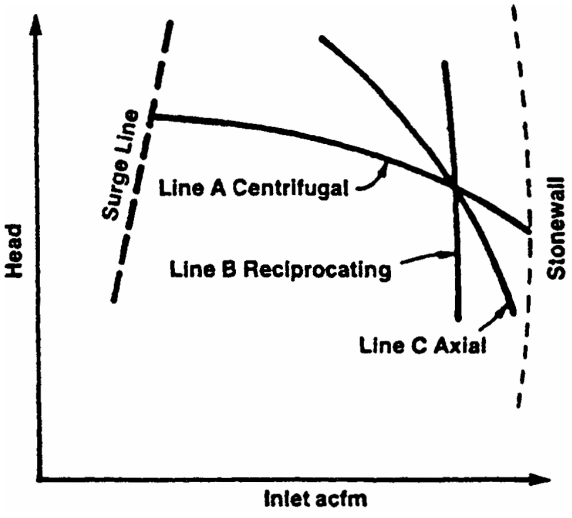


Figure 10.6: Compressor Head

10.1.12 Centrifugal Compressor Controls

Compress performance control may be effected by:

1. change of speed
2. throttling at the discharge
3. throttling at the suction
4. Guide Vane control
5. by pass control

Figure 10.5 shows a typical range of control and the effect of some of the above variables.

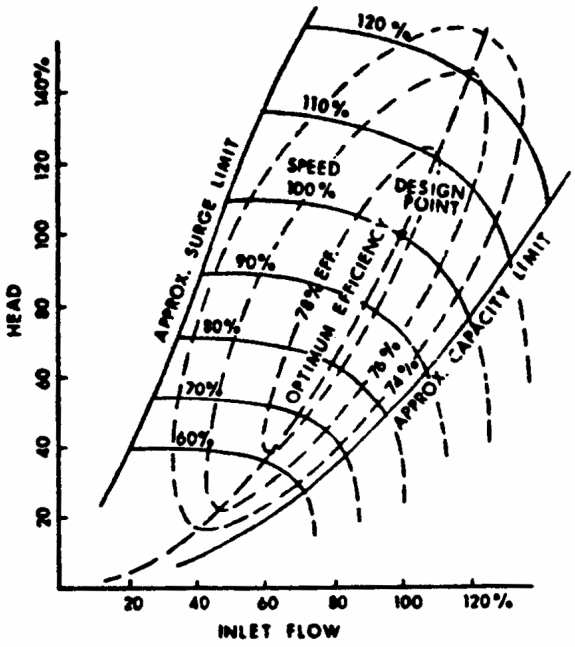


Figure 10.7: Typical Performance Map of Centrifugal Compressors

Speed changes the head of the compressor without much change in capacity. Discharge and suction throttling also changes the head but this time the system conditions.

Guide Vane control changes the shape of the performance curve and it can change significantly the surging condition.

By pass control is to prevent surging.

Surging is one of the most undesirable features of a centrifugal compressor. This occurs at a flow below which the compressor does not operate stably. The condition usually occurs when the flow is to the left of the maximum head point shown in Figure 10.7.

Figures 5.3 and 5.4 in the GPSA Data Book show how power and head are affected by rpm.

10.1.13 Specifying Centrifugal Compressors

There are also several items that should be specified for a centrifugal compressor. These include the following:

Table 10.4: Specification for Centrifugal Compressors

	Specified By
The maximum capacity required	Operator
The minimum capacity the machine is expected to operate at	Operator
The gas being compressed	Operator
The suction pressure	Operator
The discharge pressure	Operator
The driver type and speed	Operator
Operating curve showing head vs flow	Manufacturer
Number of stages	Manufacturer
Coupling to driver	Manufacturer
Compressor discharge temperature	Manufacturer
Case materials	Manufacturer
Rotating Element	Manufacturer
Control System	Operator

Considerably more data on the driver has to be supplied if the unit is to be a package of compressor and driver. Also the specification API 617, which covers centrifugal compressors, should be made part of the overall specifications.

10.1.14 Trouble Shooting Centrifugal Compressors

Operational troubles occurring in service may be due to a variety of causes.

If the trouble cannot be traced to adverse gas flow conditions or liquid "slugs" present in the system, Table 10.5 can be used as a guide for trouble shooting frequently encountered problems.

Careless operation and maintenance needs little comment. Lack of proper care of any machine is bound to result in a succession of minor troubles eventually leading to a major breakdown.

10.2 Expansion of Gases

There are many processes that use an expander as part of the process. The process is generally considered adiabatic and is the reverse of compression. The power generated is found from the equation:

Equation 10-24

$$Power = (H_1 - H_2)E$$

where

$H_1 - H_2$ = the enthalpy change of the stream assuming that the fluid is expanded from 1 to 2.

E = the efficiency which about 80 to 85% is usually run in the expander itself. Also then in the machine it is driving, such as a centrifugal compressor, efficiency must be taken into account.

The machines used are similar to steam or gas turbines but run at very high speeds. A good reference is an articles by J.S. Sweargin.

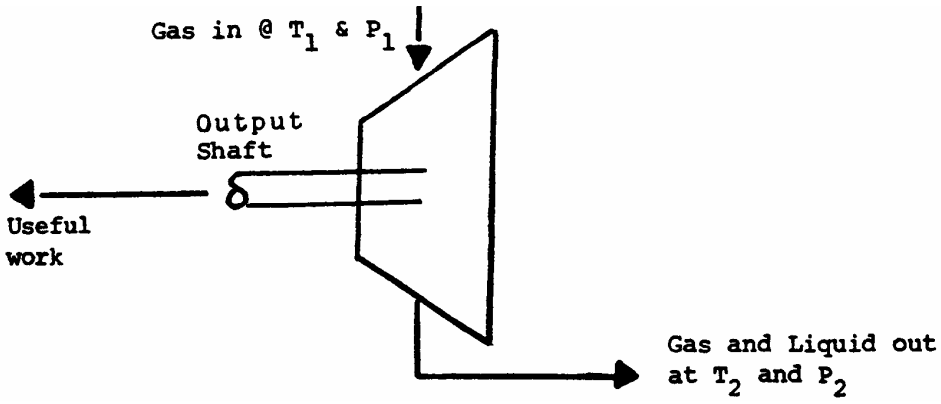


Figure 10.8: Simple Expander

Conditions for an expansion are generally set as follows: P_1 and T_1 are set by process conditions. P_2 and T_2 are set by conditions for liquid recovery and overall horsepower requirements. Generally P_2 is set and T_2 calculated and thus the expected liquid condensation and power out of the machine calculated.

Table 10.5: Probable Causes of Centrifugal Compressor Trouble

Trouble	Probable Causes
Low discharge pressure	1. Compressor not up to speed.
	2. Excessive compressor inlet temperature.
	3. Low inlet pressure.
	4. Leak in discharge piping.
	5. Excessive system demand from compressor.
Compressor surge	1. Inadequate flow through the compressor.
	2. Change in system resistance due to obstruction in the discharge piping or improper valve position.
	3. Deposit build-up on rotor or diffusers restricting gas flow.
Low lube oil pressure	1. Faulty lube oil pressure gage or switch.
	2. Low level in oil reservoir.
	3. Oil pump suction plugged.
	4. Leak in oil pump suction piping.
	5. Clogged oil strainers or filters.
	6. Failure of both main and auxiliary oil pumps
	7. Operation at a low speed without the auxiliary oil pump running (if main oil pump is shaft driven).
	8. Relief valve improperly set or stuck open.
	9. Leaks in the oil system.
	10. Incorrect pressure control valve setting operation.

Trouble	Probable Causes
High bearing oil temperature Note: Lube oil temperature leaving bearings should never be permitted to exceed 180°F.	1. Inadequate or restricted flow of lube oil to bearings.
	2. Poor conditions of lube oil or dirt or gummy deposits in bearings.
	3. Inadequate cooling water flow lube oil cooler.
	4. Fouled lube oil cooler.
	5. Wiped bearing.
	6. High oil viscosity.
	7. Excessive vibration.
	8. Water in lube oil.
	9. Rough journal surface.
Excessive vibration Note: Vibration may be transmitted from the coupled machine. To localize vibration, disconnect coupling and operate driver alone. This should help to indicate whether driver or driven machine is causing vibration.	1. Improperly assembled parts.
	2. Loose or broken bolting.
	3. Piping strajn.
	4. Shaft misalignment.
	5. Worn or damaged coupling.
	6. Dry coupling (if continuously lubricated type is used).
	7. Warped shaft caused by uneven heating or cooling.
	8. Damaged rotor or bent shaft.
	9. Unbalanced rotor or warped shaft due to severe rubbing.
	10. Uneven build-up of deposits on rotor wheels, causing unbalance.
	11. Excessive bearing clearance.
	12. Loose wheel(s) (rare case).
	13. Operating at or near critical

	11. Bearing lube oil orifices missing or plugged.
Shaft misalignment	1. Piping strain.
	2. Warped bedplate, compressor or driver.
	3. Warped foundation.
	4. Loose or broken foundation bolts.
	5. Defective grouting.

	speed.
	14. Operating in surge region.
	15. Liquid "slugs" striking wheels.
	16. Excessive vibration of adjacent machinery (sympathetic vibration).
Water in lube oil	1. Condensation in oil reservoir.
	2. Leak in lube oil cooler tubes or tube-sheet.

Courtesy of Ingersoll-Rand Co.

A very simplified method to give a order of magnitude of the power out of any expansion is to use the Mollier Diagram for methane and assume a constant entropy expansion from P_1T_1 to P_2 (and T_2). This will give you $H_1 - H_2$ in equation (24). Then assuming 80% efficiency, the approximate power available for driving the expander compressor is found.

A more accurate method is to find the amount of liquid that is condensed and thus the expected liquid recovery in a process, a trial and error type of calculation has to be done in which the outlet temperature of the expander (T_2) must be assumed if the outlet pressure is given.

Generally the method is as follows:

Given P_1, T_1 and P_2 assume 3 values for T_2 using values for T_2 less than that found from a methane Mollier diagram. At these three P_2T_2', P_2T_2'', and P_2T_2''' conditions, calculate the equilibrium conditions for liquid and vapour present, the enthalpy and entropy at each condition and plot the entropy and the change in enthalpy from 1. (i.e. $H_1 - H_2$ or ΔH). All as shown on the figure below.

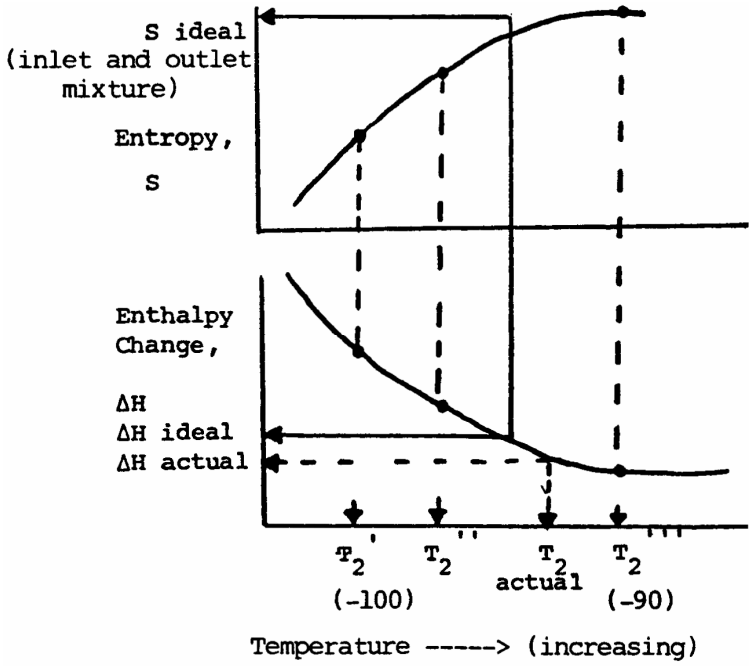


Figure 10.9: Entropy and Enthalpy Plot

Find entropy at the outlet conditions which is to equal the inlet condition entropy and then the corresponding H which is the ideal H. Assume an efficiency and calculate the actual H and thus find the actual T_2. Run a flash calculation at this temperature to see how much liquid condenses so as to have an estimate of the liquid recovery.

Pages 5-11, 5-12 and 5-13 of the GPSA Manual go into more detail of this calculation and give an example calculation.

Figure 10.10 shows the steps of the calculation more explicitly.

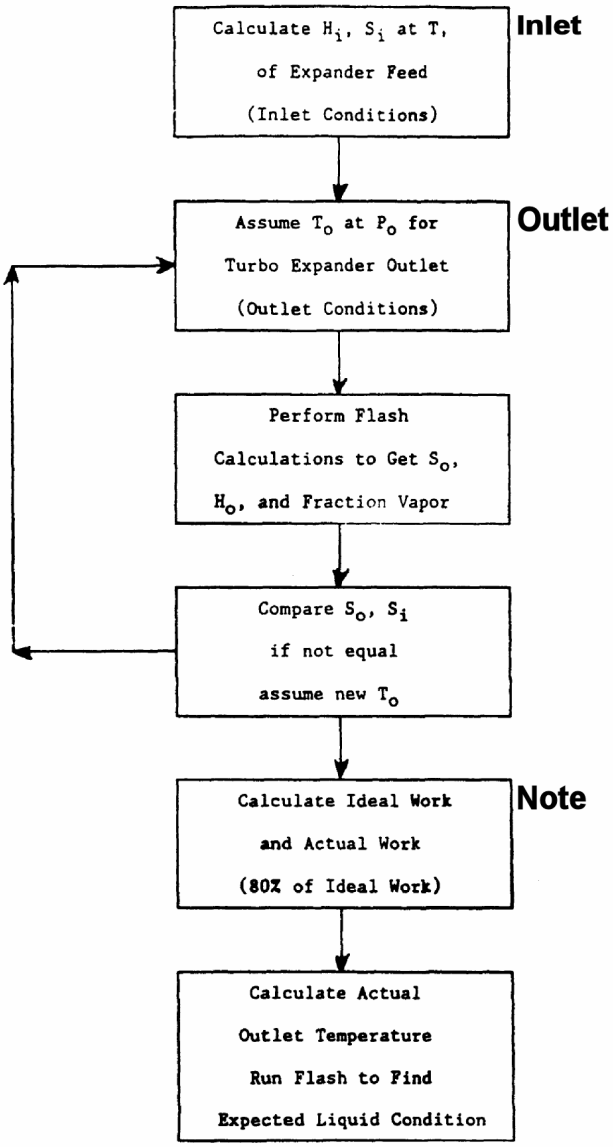


Figure 10.10: Finding the Amount of Condensed Liquid

10.3 Refrigeration

We are concerned mostly with the general theory of operation of the refrigeration system and not all the mechanical details.

10.3.1 General

To start let us consider the following simple system (Figure 10.11).

The process is as follows:

Liquid refrigerant is vaporized in the evaporator to remove heat. It is recompressed by the compressor to high pressure, then condensed at the high pressure and collected in the surge drum. The liquid so formed is then throttled through the throttling valve into the chiller where all the liquid is evaporated and the cycle is repeated.

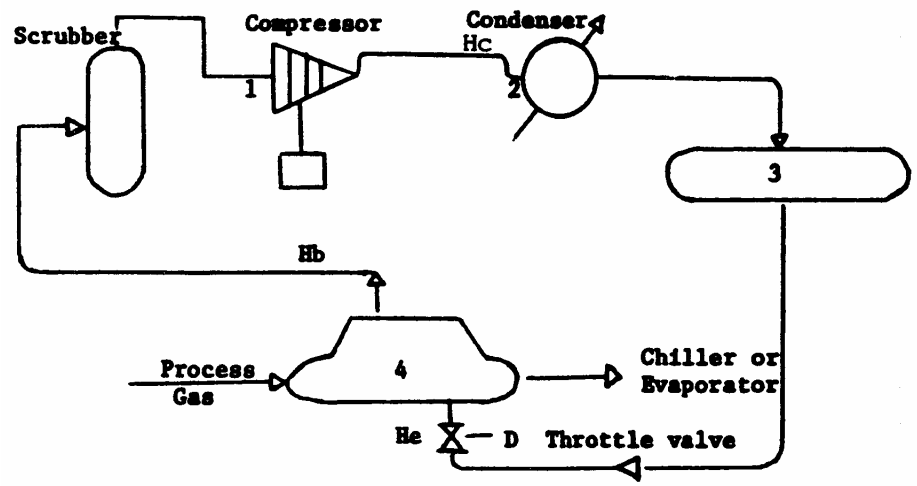


Figure 10.11: Typical Refrigeration System

The prime principle of a refrigeration machine is the absorbing of heat at a low temperature in the evaporator and rejecting this heat at a higher temperature in the condenser. It also depends on the Joule Thompson effect of constant enthalpy expansion.

Basically the cycle can be broken down into four steps as discussed above i.e. Expansion, Evaporation, Compression and Condensation and as shown on Fig. 10.12.

In Figure 10.11 let us define the heat conditions as follows:

Hb = heat content of refrigerant leaving chiller.

Hc = heat content of refrigerant leaving compressor.

He = heat content of refrigerant entering the chiller.

Also we define tons of refrigeration = amount of heat removed such that

1 ton = 200 BTU/min or 12,000 BTU/hr

In metric units we measure the KW of refrigeration and the brake power per KW of refrigeration.

The heat abstracted is

Hb - He

The work done is (assuming adiabatic compression) the change in enthalpy

Hc - Hb

The coefficient of performance β is defined as follows:

Equation 10-25

$$\beta = \frac{Hb - He}{Hc - Hb}$$

Now to find HP required:

Equation 10-26

$$HP\ per\ ton\ of\ refrigeration = \frac{4.713}{\beta}$$

This is theoretical HP of course. To find actual amount we have to use an efficiency factor. Take about 75-80%. Thus actual HP is

Equation 10-27

$$\frac{Theoretical\ HP}{0.80}$$

It is also possible to find theoretical HP by finding

(Hc - Hb) (W) and converting BTU's to HP.

W = lbs of refrigerant circulated/unit of time.

10.3.2 Thermodynamics

To understand the refrigeration process let us examine a pressure enthalpy diagram and point out the conditions that occur in the cycle.

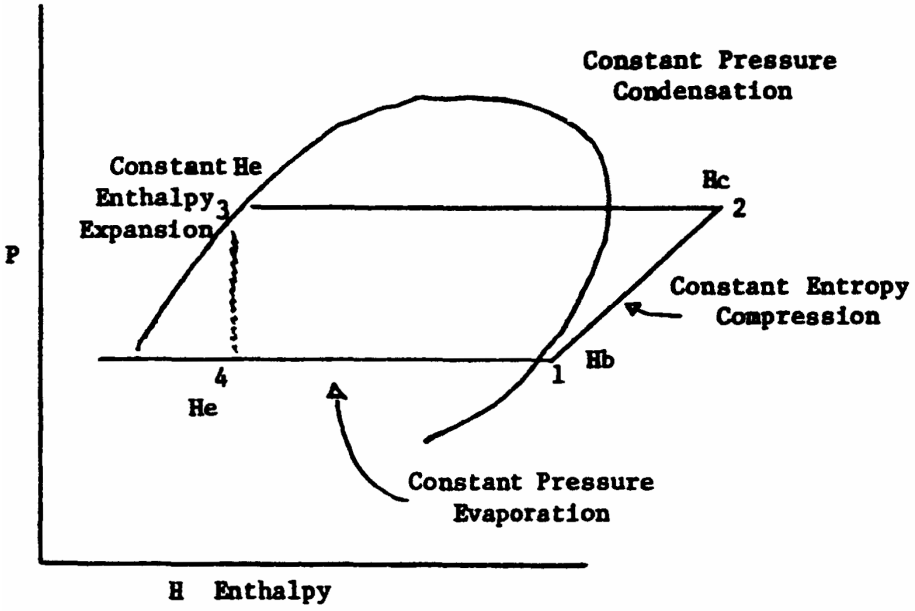


Figure 10.12: Refrigeration Thermodynamic Cycle

At Point 1 we have a refrigerant vapour at low temperature and ready to be compressed adiabatically (enthalpy Hb). At Point 2 we have a compressed gas at the outlet of the compressor (enthalpy Hc). It is then cooled and condensed to enthalpy "He". It is then throttled (i.e. expanded) at constant enthalpy to Point 4. Finally the refrigerant is evaporated and returned to the compressor suction.

There are several modifications to the simple refrigeration system that can reduce the horsepower required in the system. The most common system is to use stages as described below.

10.3.2.1 Refrigeration Stages

Refrigeration systems utilizing one, two, three, or four stages of compression have been successfully operated in various services. The number of levels of refrigeration generally depends upon the number of compression stages required, interstage heat loads, economics, and the type of compression.

One-Stage Refrigeration System

A typical one-stage refrigeration system is shown in Fig. 10.13 where the data are for pure propane refrigerant.

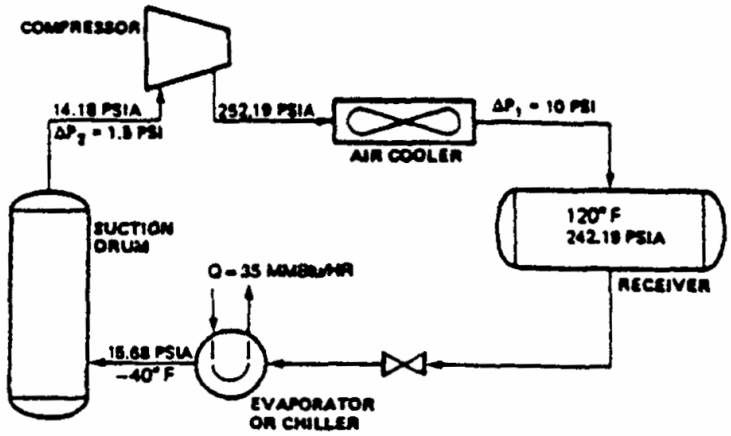


Figure 10.13: One-Stage Refrigeration System

Two-Stage Refrigeration System

Savings in the 20% range can often be achieved with a two-stage refrigeration system and interstage flash economizer. Additional savings can be realized by removing process heat at the interstage level rather than at the low stage level. A typical two-stage system with an intermediate load is shown in Fig. 10.14 with data for pure propane.

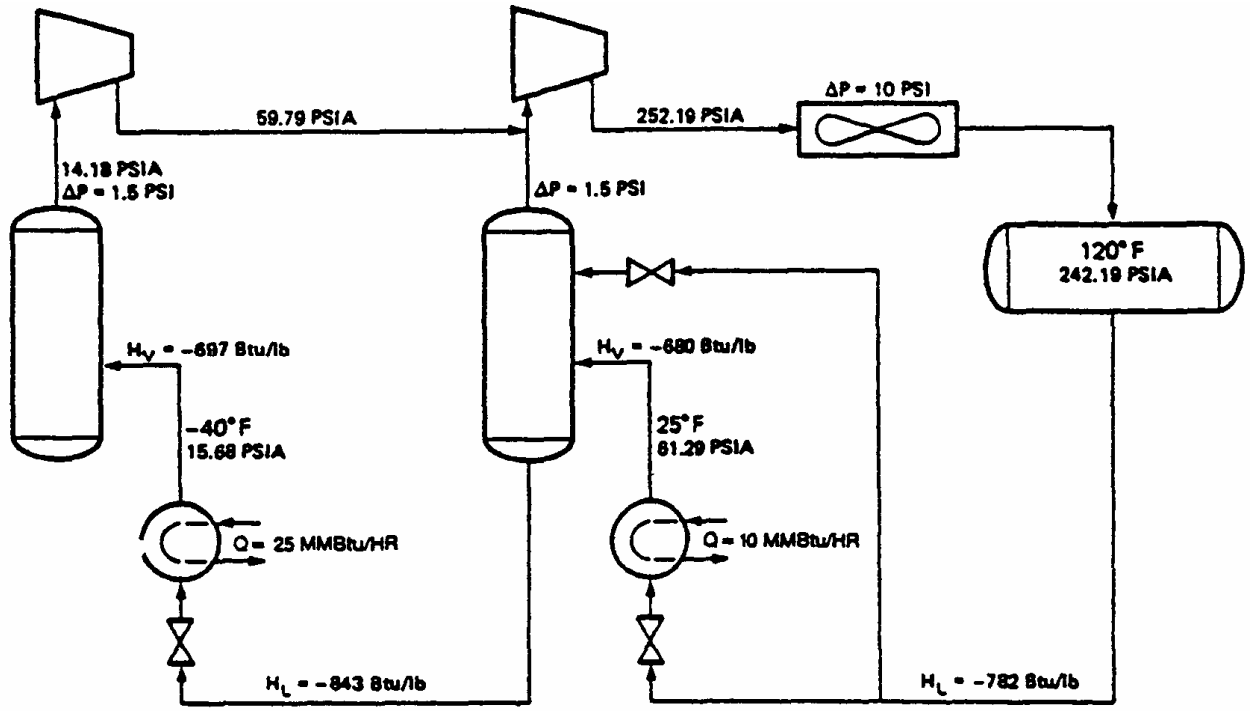


Figure 10.14: Two-Stage Refrigeration System

Three-Stage Refrigeration System

Additional horsepower savings can be achieved by using a three-stage compression system. As with a two-stage system, flash economization and/or an intermediate heat loads can be used. The savings, while not as dramatic as the two stage versus one-stage, can still be significant enough to justify the additional equipment. A typical three stage propane system is shown in Fig. 10.15.

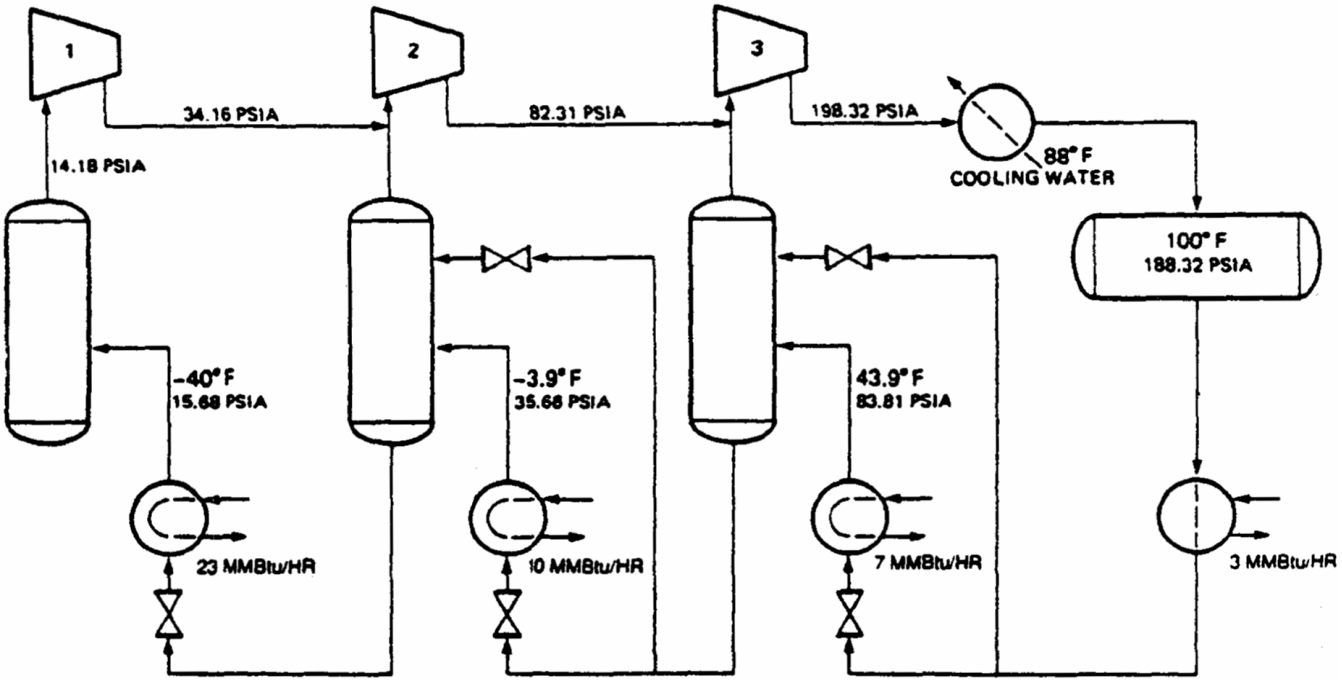


Figure 10.15: Three-Stage Refrigeration System

Refrigeration System Configuration

Energy consumption is frequently reduced as the number of stages is increased. For a propane refrigeration system, Table 10.6 illustrates the effect of interstages without using refrigeration at intermediate levels. However, the installation cost of such refrigeration systems increases as the number of stages increases. The optimum overall cost will be a function of the specific system and has to be determined for a set of economic criteria.

Table 10.6: Effect of Staging on a Propane Refrigeration System

	Stages, n		
	1	2	3
Refrigeration Duty, MMBtu/hr	1.0	1.0	1.0
Refrigeration Temperature, ºF	- 40	- 40	- 40
Refrigerant Condensing Temperature, ºF	100	100	100
Compression Requirements, hp	292	236	224
Reduction in hp, %	Base	19.2	23.3
Condenser Duty, MMBtu/hr	1.743	1.600	1.575
Change in condenser duty, %	Base	-8.2	-9.6

10.3.3 Refrigerants Used

The common refrigerant used in gas processing is propane as it is normally readily available in a plant. However propylene, freon and ammonia can also be used in the same service. For colder temperatures in LNG cycles ethylene and methane are the normal refrigerants.

10.3.4 Power Required

The GPSA Data Book gives a comparison of the atrinonia, propylene, propane and freon in regards to HP/ton of refrigeration (or Brake power/KW of refrigeration), etc. on Fig. 5.18.

To obtain a more accurate power requirement Figure 10.11, Figure 10.12 and Figure 10.13 can be used to calculate this value and the condenser duty depending on the evaporator temperature and the refrigerant condensing temperature and the number of stages.

10.3.5 Equipment Required

a) Compressors - Both reciprocating compressors and centrifugal used. Centrifugal are preferred because of lower maintenance generally not used when HP requirements are below 1,000.

b) Condensers - The condenser must remove all the heat taken out of the evaporator plus the heat of compression. The condenser can be either water cooled or air cooled.

c) Water Cooled - These are standard, well-designed shell and tube exchangers. They have the advantage of making the system have less HP than a system with an air cooled condenser. Figure 10.14 shows the difference of HP required with various condensing temperatures for a propane refrigerant system against various evaporation temperatures.

d) Air Cooled - These are standard air cooled condensers. The lowest condensing temperature that probably can be expected is about 20ºF above the design ambient temperature and thus the system requires more HP.

e) Economizer - This is an intermediate flash vessel that gets rid of some of the vapours that would be formed on direct throttling. Two stages are often used when more than two compression stages are used.

f) Throttling Valve - This is a standard control valve mostly operated on level control from the evaporator of chiller.

g) Evaporator (or Chiller) - A standard U-tube heat exchanger in which the refrigerant is normally on the outside of the tubes is used. A good disengaging space should be provided in the top of the refrigerant side to prevent liquid entrainment.

h) Suction Scrubber - This is a large vessel complete with entrainment separators, automatic liquid level dumps, etc. It is there to prevent any liquid carryover into the compressors.

i) Auxiliaries - Such units as refrigerant driers, oil removal facilities, purges, etc., are also generally provided.

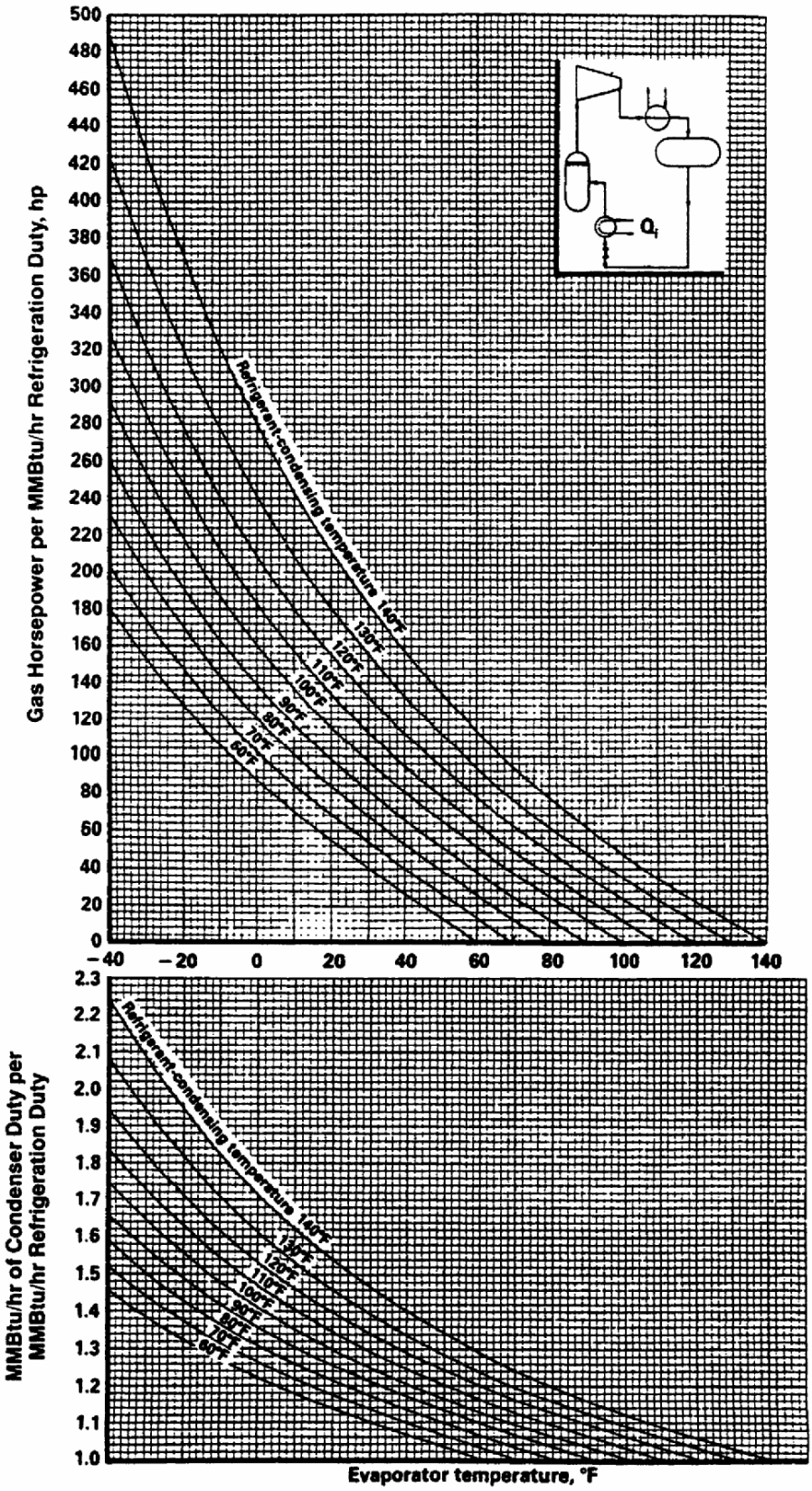


Figure 10.16: Single-Stage Propane Refrigeration System

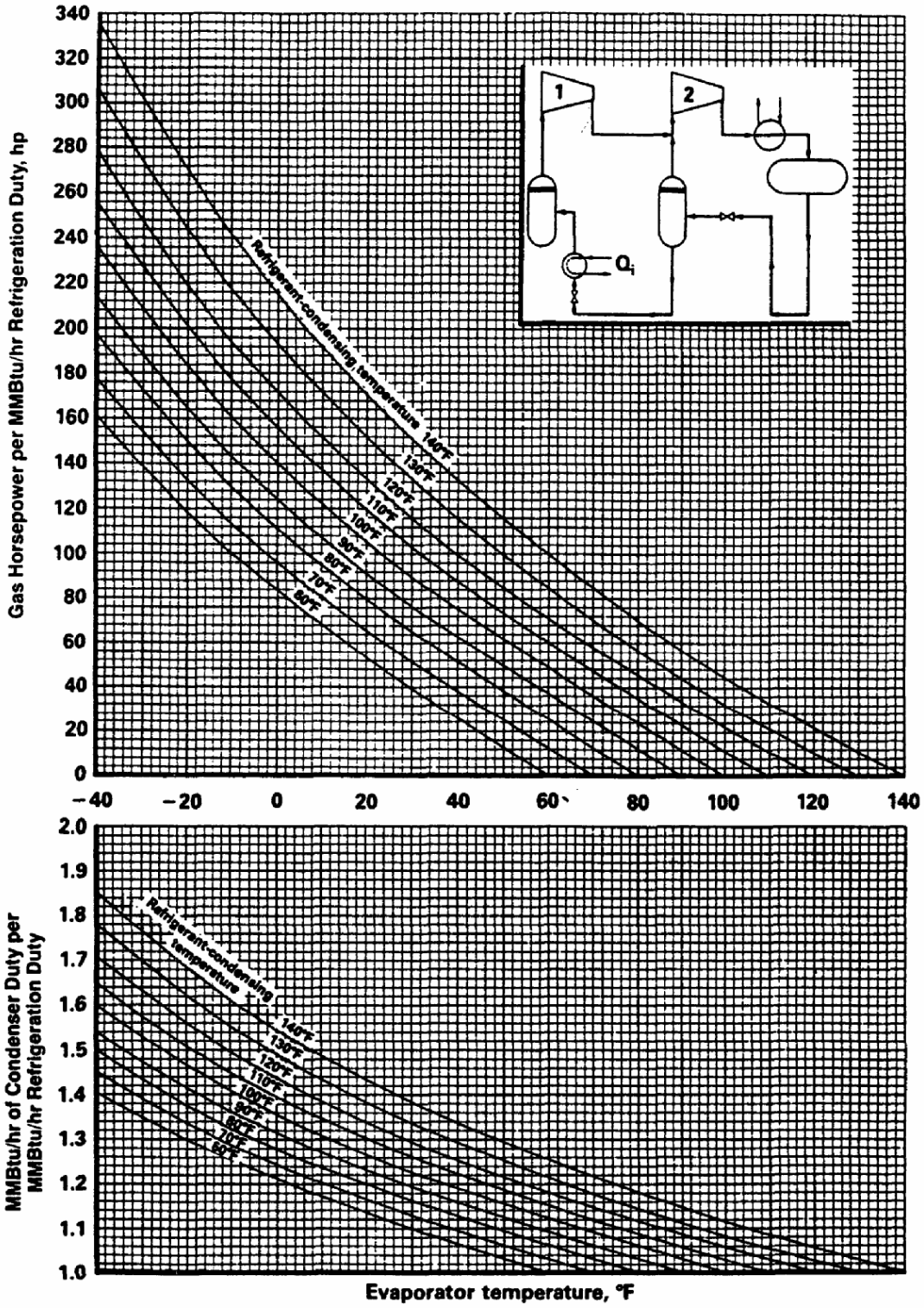


Figure 10.17: Two-Stage Propane Refrigeration System

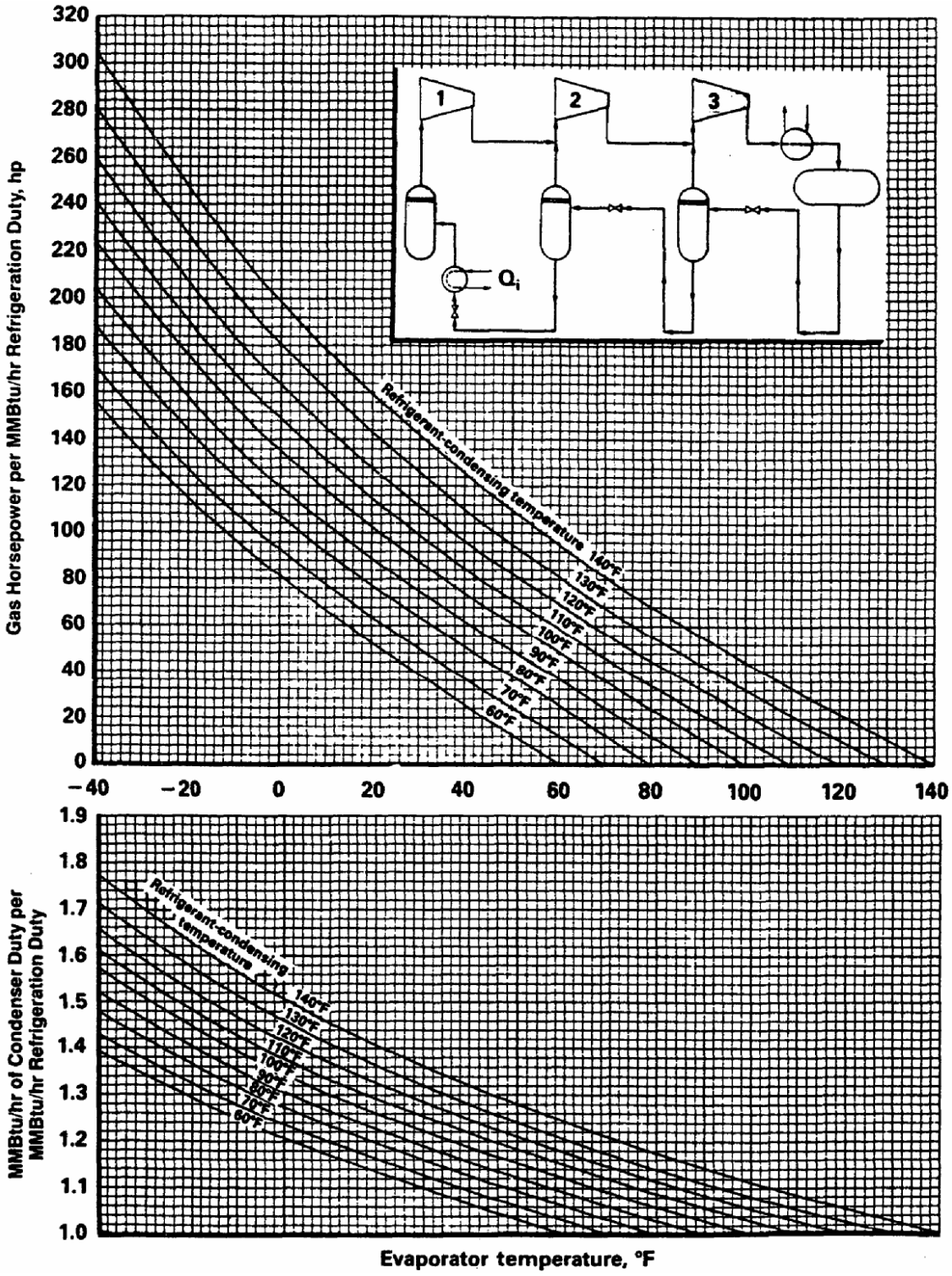


Figure 10.18: Three-Stage Propane Refrigeration System

10.3.6 Conditions for a Chilling System

Evaporator Temperature

This should be set at about 10ºF lower than the process outlet temperature. This sets the evaporator pressure and thus the suction pressure to the compressor.

Condenser Refrigerant Outlet Temperature

This is set by the type of cooling medium. With water it is about 10ºF above the water outlet temperature and thus runs about 90ºF to 100ºF. With an air cooled condenser, again the propane will be 10ºF above the outlet temperature of the aerial cooler outlet temperature which will probably run about 100ºF to 110ºF. Thus a propane temperature of 110ºF to 120ºF will be obtained. This information sets the compressor discharge pressure. From the above data the compressor HP can be calculated. The following example shows how a typical system is calculated:

10.3.7 Refrigeration System Checklist

Indication	Causes
High Compressor.	Check accumulator temperature.
Discharge Pressure.	If the accumulator temperature is high, check: 1. Condenser operation for fouling. 2. High air or water temperature. 3. Low fan speed or pitch. 4. Low water circulation. If condensing temperature is normal, check for: 1. Non-condensables in refrigerant. 2. Restriction in system which is creating pressure drop.
High Process Temperature.	Check refrigerant temperature from chiller. If refrigerant temperature is high and approach temperature on chiller is normal, check: 1. Chiller pressure. 2. Refrigerant composition for heavy ends contamination. 3. Refrigerant circulation or kettle level (possible inadequate flow resulting in superheating of refrigerant). 4. Process overload of refrigerant system. If refrigerant temperature is normal, and approach to process temperature is high, check: 1. Fouling on refrigerant side (lube oil or moisture). 2. Fouling on process side (wax or hydrates). 3. Process overload of chiller capacity.
Inadequate Compressor Capacity	Check: 1. Process overload of refrigerant system. 2. Premature opening of hot gas bypass. 3. Compressor valve failure. 4. Compressor suction pressure restriction. 5. Low compressor speed.
Inadequate Refrigerant Flow to Economizer or Chiller	Check: 1. Low accumulator level. 2. Expansion valve capacity. 3. Chiller or economizer level control malfunction. 4. Restriction in refrigerant flow (hydrates or ice).

Example Problem 10.3

Calculate the HP required for refrigeration to chill 100 MMSCFD of a gas at 937 psig of the composition listed below from 40ºF to 0ºF using propane as a refrigerant. The evaporator temperature is -10ºF and the propane condenser operates to condense the propane to a liquid at 95ºF.

Composition	Mole %
N_2	0.00
C_1	95.62
C_2	3.01
C_3	0.67
iC_4	0.18
nC_4	0.20
iC_5	0.09
nC_5	0.09
C_6	0.14

Solution

First find the amount of heat to be removed in the chillers. This is found by enthalpy calculations.

This calculation is shown in Example Problem 2 in Chapter 2. The duty is equal to 5,727,028 BTU/HR or $\frac{5{,}727.028}{12000} = 418$ tons of refrigeration.

In metric this equals 1679 KW.

Next calculate the refrigeration HP, finding the coefficient of performance:

$$\frac{H_b - H_e}{H_c - H_b}$$

For this condition, using Propane Mollier Diagram in GPSA Data Book:

H_b at -10ºF and saturated vapour = -690 BTU/lb

H_e at 95ºF and saturated liquid = -800 at (175 psia-condensing pressure)

H_c at 118ºF and 175 psia = -652

$$\beta = \frac{-690 - (-800)}{-652 - (-690)} = 2.90$$

HP per ton of refrigeration

$$\frac{4.713}{\beta} = \frac{4.713}{2.90} = 1.62$$

This is a theoretical value. Divide by efficiency of 80% to get brake HP

$$\frac{1.62}{.80} = 2.02\ HP/ton$$

Compare this with value obtained from Fig. 5-14 of the GPSA Data Book.

2.03 HP/ton for conditions:

HP required using calculated HP/ton = 2.02 (481) = 971.6

Therefore use 1000 HP.

In metric units the Brake Power/KW of refrigeration is 0.43 at -23.3ºC from Fig. 5.18 in metric GPSA Data Book. Therefore the Brake Power required: 1679.2 (.43) = 722 KW.

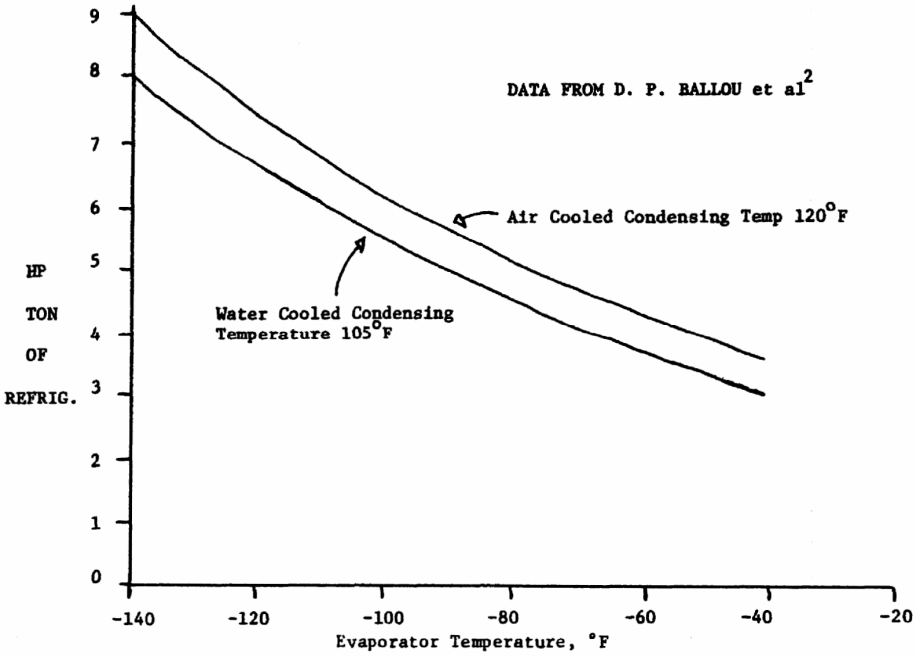


Figure 10.19: HP Required for Different Condensing System

10.4 Example Problems

Example Problem 10.4

A problem has arisen in an absorption plant that has an average flow of 600 MMSCFD. Too much ethane is being absorbed such that at maximum recoveries of propane of 85% of the stream it is necessary to flare approximately 4 MMSCFD of a mixture of 10% C_1 and 90% C_2. This is not allowed by governmental regulations. The flaring can be stopped by cutting back the absorption rate so that only 83% of the propane is recovered. Instead it is proposed to install a compressor to reinject the gas into the sales gas line. The capital cost of the compressor will be $650/HP installed and it will cost to operate $10/HP/year including fuel. The value of gas to the plant is $1.05/MMBTU. The value of propane is $7.00/barrel when sent out of the plant. The ethane gas is available at dethanizer overheat at 400 psig and the sales gas line operates at 925 psig. The pressure drop in any piping and controls to the sales gas line will be 15 psi. The gas composition to the plant is:

Composition	Mole %
C_1	93.0
C_2	4.8
C_3	1.4
iC_4	0.3
nC_4	0.5

The ethane overhead stream is colled to 60ºF before it is compressed. Assume an overall efficiency to the compressor of 85%. Determine the payout in years before taxes on the installation of this facility to see if it is worthwhile. Also determine the DCF rate of return if the installation has a 25 year life.

Example Problem 10.5

For a special deep chilling process for lpg recovery it is planned to use ethylene as the refrigerant. The gas in the process is to be chilled to -110ºF using ethylene with a 10ºF approach in the chiller.

The ethylene is condensed by propane to 0ºF in the ethylene compressor condenser.

If the ethylene compressor has to take out 600 tons of refrigeration, calculate the HP required by the ethylene refrigeration machine. Assume the efficiency of the machine is 75%.

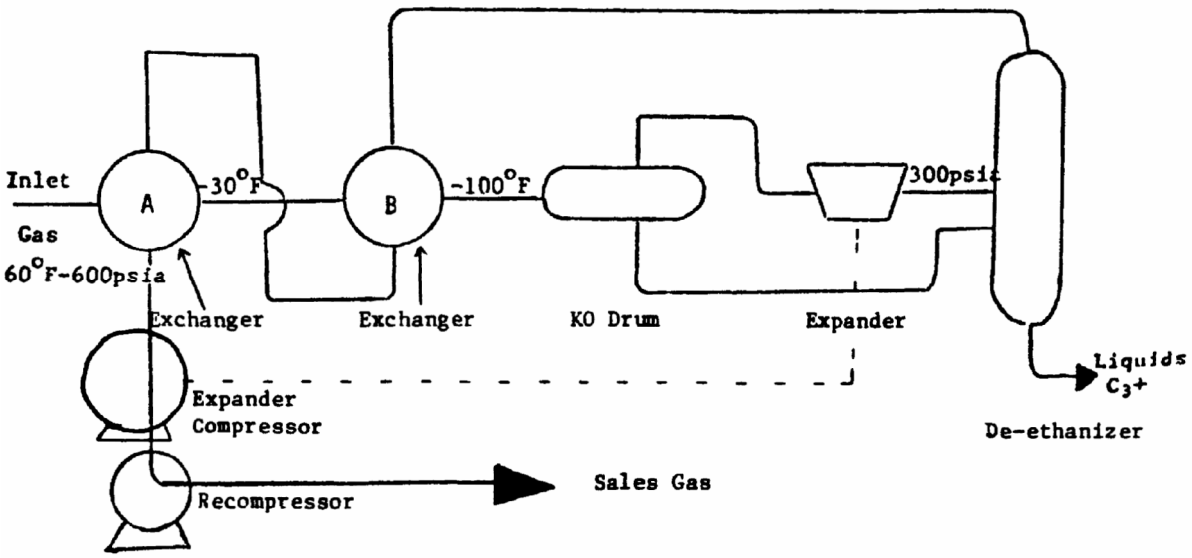
Also calculate the HP (assume same efficiency) for the propane compressor if it is used to cool the gas flow at 200 MMSCFD (0.6 SpG at 1000 psig) from 40ºF to 0ºF as well as condense the ethylene.

Example Problem 10.6

A plant that is on a sales gas line is designed to recover lpg's (C_3^+) uses the expander process. The design volume of gas is 400 MMSCFD. The inlet gas analysis is:

Composition	Mole %
C_1	92.00
C_2	5.40
C_3	1.81
iC_4	0.29
nC_4	0.40
iC_5	0.10
Average MW	17.5

A simple flow sheet of the plant is shown below.



Exchanger B ruptures and is out of service for 4 months. You are to decide what is the best way to continue processing the gas from the following choices:

1. Bypass exchanger B on both sides and handle 400 MMSCFD in the expander. In this case the outlet process gas temperature from A is reduced to -40ºF because A has colder stripped gas entering it.
2. Bypass B and the expander and reduce the pressure to 300 psia by means of a throttle valve. The outlet of exchanger A is reduced to -80ºF in this case.

Assume that the reduction in flow is proportional to loss in HP from the expander being shut down, that is:

$$400\ MMSCFD\left(\frac{HP\ to\ system\ \ from\ Expander}{\mathrm{Re}\,compressor\ HP + HP\ to\ system\ \ from\ Compressor}\right)$$

The recompressor HP is 24,000.

The expander system is 60% overall efficient.

Assume 5 psi drop per exchanger.

Assume pure methane for all expander enthalpy, entropy calculations.

Assume all C_3^+ that is liquefied is recovered.

Example Problem 10.7

One of the schemes for moving gas from the Arctic involves chilling the gas initially to remove the liquids and then keeping it cold so not to melt the permafrost.

Please calculate the approximate HP required at a compressor station that handles 5 MMSCFD of gas (measured at 14.4 psia and 60ºF) and the following takes place.

- Gas is compressed from 400 psia to 800 psia. The suction temperature is +25ºF.
- The gas is chilled from +80ºF to -5ºF so as to keep the gas cool enough to reach the next station at not greater than +25ºF.

Compare this with the case in which no chilling is put in the station and the gas is compressed at a suction temperature of +75ºF (summer conditions) from 400 psia to 800 psia.

Assumptions

a) Gas properties

MW = 18

c_P/c_V = 1.26

b) Enthalpy difference of cooling gas from 80ºF to -5ºF at 800 psia is 43 BTU's/lb.

c) Use propane as the chilling medium. Assume a 5ºF approach in the chiller and assume a 95ºF propane liquid condenser temperature.

d) Assume pipeline compressors are 85% efficient and that reciprocating compressors are used.

References

3. J. S. Sweargin, Hydrocarbon Processing, April 1970, P. 97.
4. D. P. Ballou et al., Hydrocarbon Processing, June 1967, P. 119.
5. GPSA Data Book 1987 Edition Chapters 13 & iii.

11.0 Economics

- A review of the types of calculations used in justifying a project, ie. payouts, average annual rates of return.
- A look at the data required for these calculations.

11.1 Introduction

The determining of whether a project is commercially feasible is, of course, very important. There are several methods of determining whether a project or venture is economically desirable. These methods and their results are also used to decide which of many alternates are the most desirable.

11.2 Economics Indicators

a) Payout - This is the time from initial operation it takes for the project to generate enough cash to recover the investment made in the project. The cash flow is the sum of the net income after income tax plus the allowable income tax depreciation and, if allowed, the depletion allowance.

b) Discount Cash Flow Rate of Return - This is the interest rate which the actual investment is earning over the life of the project.

c) Number of Times Investment Return - This is the number of times that the earnings recover the investment. It is found by dividing the cash flow for the life of the project by the investment made.

d) Present Worth - This is the present day value of all future earnings. This is found by discounting the future earnings at a prescribed discount rate. Generally both a) and b) and often c) and d) should be worked out for each project. If alternates are being considered, all should be obtained.

e) Present Value Profit Ratio - This is the present value of all earnings at prescribed discount rate divided by the capital cost. Often companies set the minimum DCF rate of return acceptable and then calculate the present worth at this rate. The measure of a project is the present worth of a project above this minimum.

f) Others - There are several economic indicators such as the equivalent rate of return, simple rate of return, etc.

11.3 What is Considered a Good Investment

This, of course, depends a great deal on the size and type of project and on the risks involved in the project. The risks involved can be outlined as follows:

a) Markets Changing - Markets for products can disappear. This is unlikely with gas and by-products, which are primarily fuel. However, if they are not priced competitively, they can be reduced drastically as has happened to Canadian natural gas in the United States. For example, sulphur may be replaced if the price is too high. Also, there is the possibility of reduced market prices. This was not so significant with gas, but recently it has become a real possibility and is very much the case with LPG's and sulphur, but not condensates. For

example, in 1967 sulphur was being sold at $50/LT f.o.b. plants and in 1971 was down as low as $4/LT. It has subsequently risen to over $150 in 1985.

b) Change of Costs - Plant costs can change, both capital and operating; these changes can generally be reasonably forecasted. However, with the shortages of labour and material, the cost of plants, pipelines, etc. go up, then with a recession they go down again as witnessed by 1983 and 1984 costs as compared to 1981 and 1982 costs. See the Nelson Index table later in this Section showing the changes with time. Inflation is generally a very big factor in capital costs. The risk in regard to plant capital costs are discussed later with the accuracy expected in cost estimating. (Also see reference #5 on this.)

c) New Technology - New plants become cheaper because of new technology. This change is not so important in gas processing as compared to petrochemical plants because the main product, gas, is often contracted for the life of the plant. Also, there has not been a major technical change in gas processing in the last ten years.

d) Government Regulations - This change is very difficult to forecast, as witnessed by the many changes in 1974, 1975 and 1981 in the petroleum industry via taxes, royalty rates, PIP grants, etc. This has become the most variable and unpredictable factor in economic evaluations.

All these factors have to be taken into account when evaluating projects. The higher the risk, the better the economic factors should be.

An article by D. Stahlberg discusses how to identify some of the risks. A formula that identifies the risk is as follows:

Equation 11-1

$$\%\ Risk = \frac{(Original\ Net\ Cost + Cost\ of\ Way\ Out - Net\ Cost\ of\ Safe\ Plan)}{Net\ Cost\ of\ Safe\ Plan - Original\ Net\ Cost} \times 100\%$$

This % risk is compared with the chance of success, and if the success chance greatly outweighs the risks, then the riskier method should be tried.

Here are some guidelines for projects.

Minimum Payouts

For gas processing projects:

Before Taxes	3 - 4 years
After Taxes	6 - 7 years

For petrochemical projects where the risks are probably greater:

Before Taxes	2 - 3 years
After Taxes	5 years

Minimum Discounted Cash Flow After Taxes

For Gas Processing Plants	11% - 15%
For Petrochemicals	20%
For Pipelines	12%

11.4 Capital Costs

There are many types of capital cost estimates. The accuracy of these vary a great deal depending on the amount of information obtained and work done by the estimator. There is a great deal of information and data on the type and scope of estimates. A whole course could be given on making proper estimates. Also, there is a professional society devoted to cost engineering, the American Association of Cost Engineers. Figure 11.1 shows the approximate accuracy of estimates depending on the status of the project. Also Figure 11.2 gives a graphical outline of some of the types of estimates and the information needed for them. Each type of estimate requires a different amount of data to obtain.

11.4.1 Preliminary Feasibility Estimate

The preliminary feasibility estimate (order of magnitude).

As a very general guide the following can be used for gas plant costs:

Type of Gas	Capital Cost - $/MMSCF
Very sour and rich	300,000 - 500,000
Medium sour	100,000 - 300,000
Sweet but rich	50,000 - 100,000
Dry and sweet	10,000 - 50,000

or you can use the method of E. Berlie.

11.4.2 Preliminary Budget Estimate

The concept or preliminary budget estimate can be done if you have the overall data on gas analysis, volumes, etc. This estimate is done by breaking the plant down into various process units and using general data for gathering pipelines, etc. The following table gives the curves that are included.

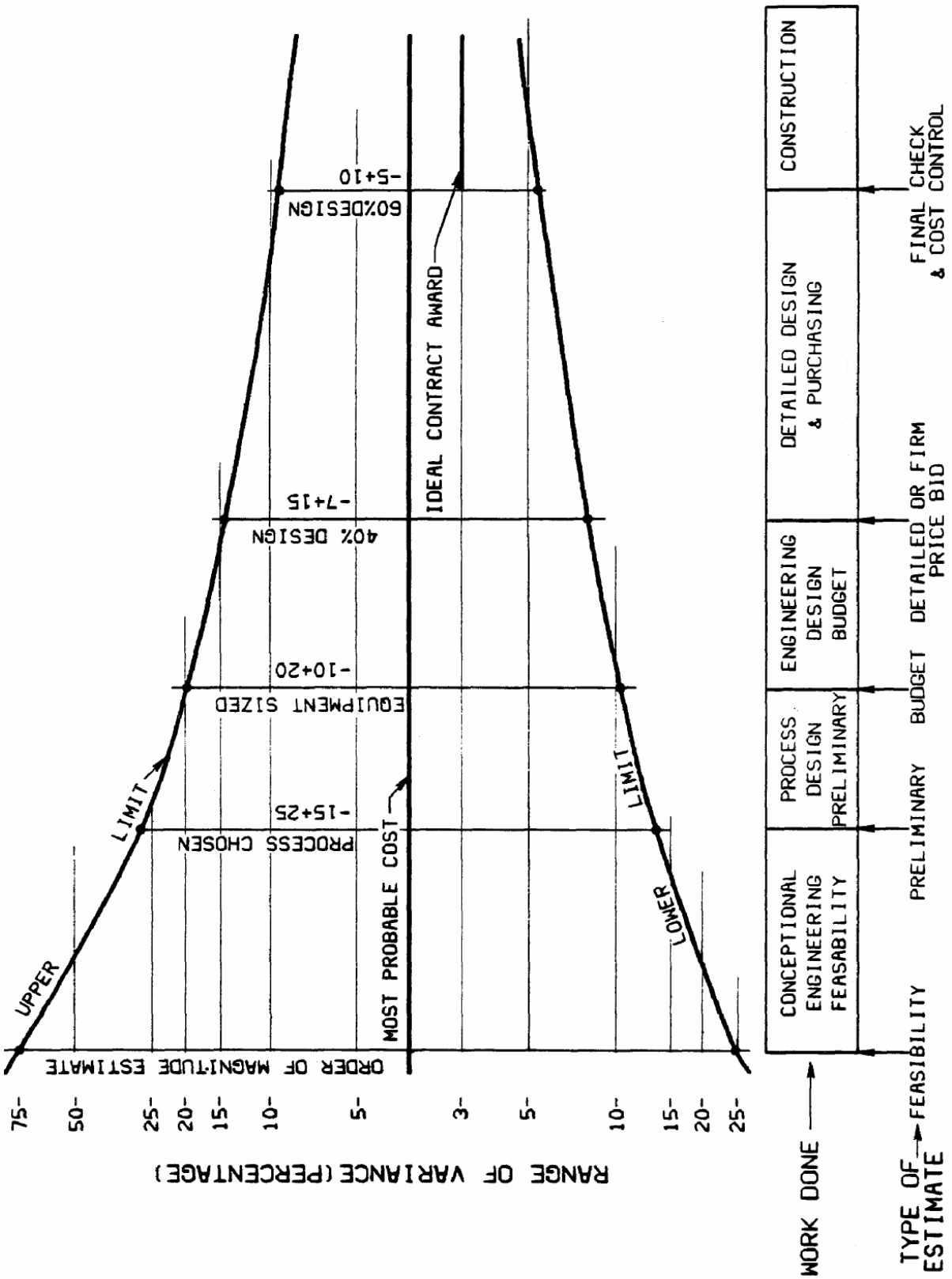


Figure 11.1: Estimate Accuracy

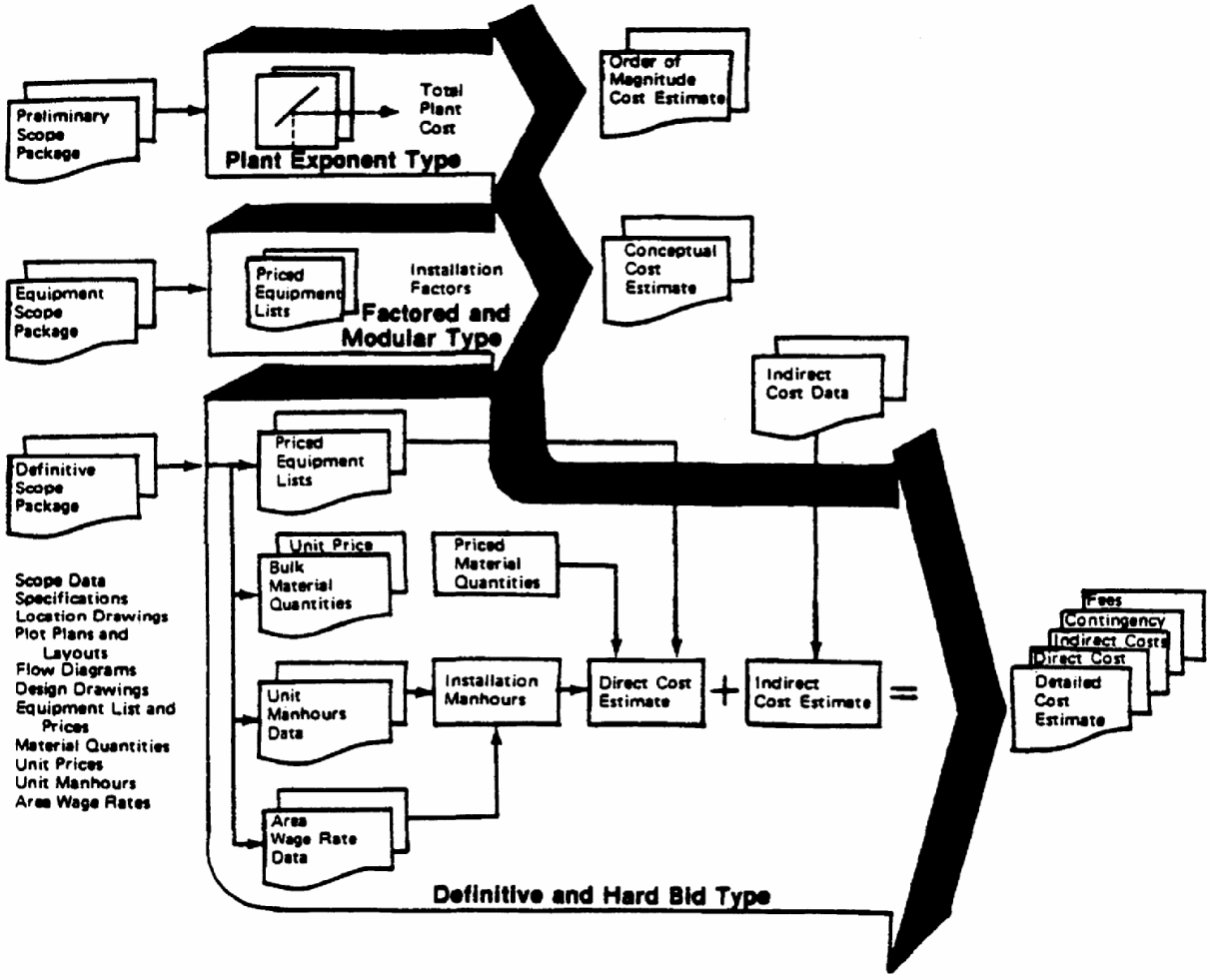


Figure 11.2: Categories of Estimates

Table 11.1: Process Unit Cost Curves

Figure No.	Process Unit	Variable
Figure 11.3	Sweetening	US gpm solvent circulated
Figure 11.4	Sulphur Recovery	LTD sulphur product
Figure 11.5	Dewpoint Control	MMSCF Feed Gas
Figure 11.6	Glycol Dehydration	MMSCF Feed Gas
Figure 11.7	Fractionation	Feed MBPD
Figure 11.8	Compressor Installation	BHP
Figure 11.9	Gathering System Cost	Diameter inch mile
Figure 11.10	Wellsite Cost	Flow MMSCFD
Figure 11.11	NGL's Facilities	Recovery of Liquids BPD

Note: This information is presented courtesy of G. E. Handwerk, Consulting Engineer, Denver, Colorado, Fish International Canada Ltd., Calgary, Alberta, Dome Petroleum Limited, Calgary, Alberta and SKM Consulting Ltd., Calgary, Alberta.

The year of information in the curves is given and should be adjusted to present day costs using an inflation index as indicated below.

The cost of various facilities (particularly plants) changes with time. A good way to update costs is to multiply the cost of the facility in a certain year by the ratio of the "Nelson" construction index of today's date and divide by the Nelson construction index of the date the plant is built. Therefore, if you know that a plant costs $10,000,000 in 1964, the cost for the same plant in 1976 would be:

$$\$10{,}000{,}000\,\frac{626.1}{257} = \$24{,}362{,}000$$

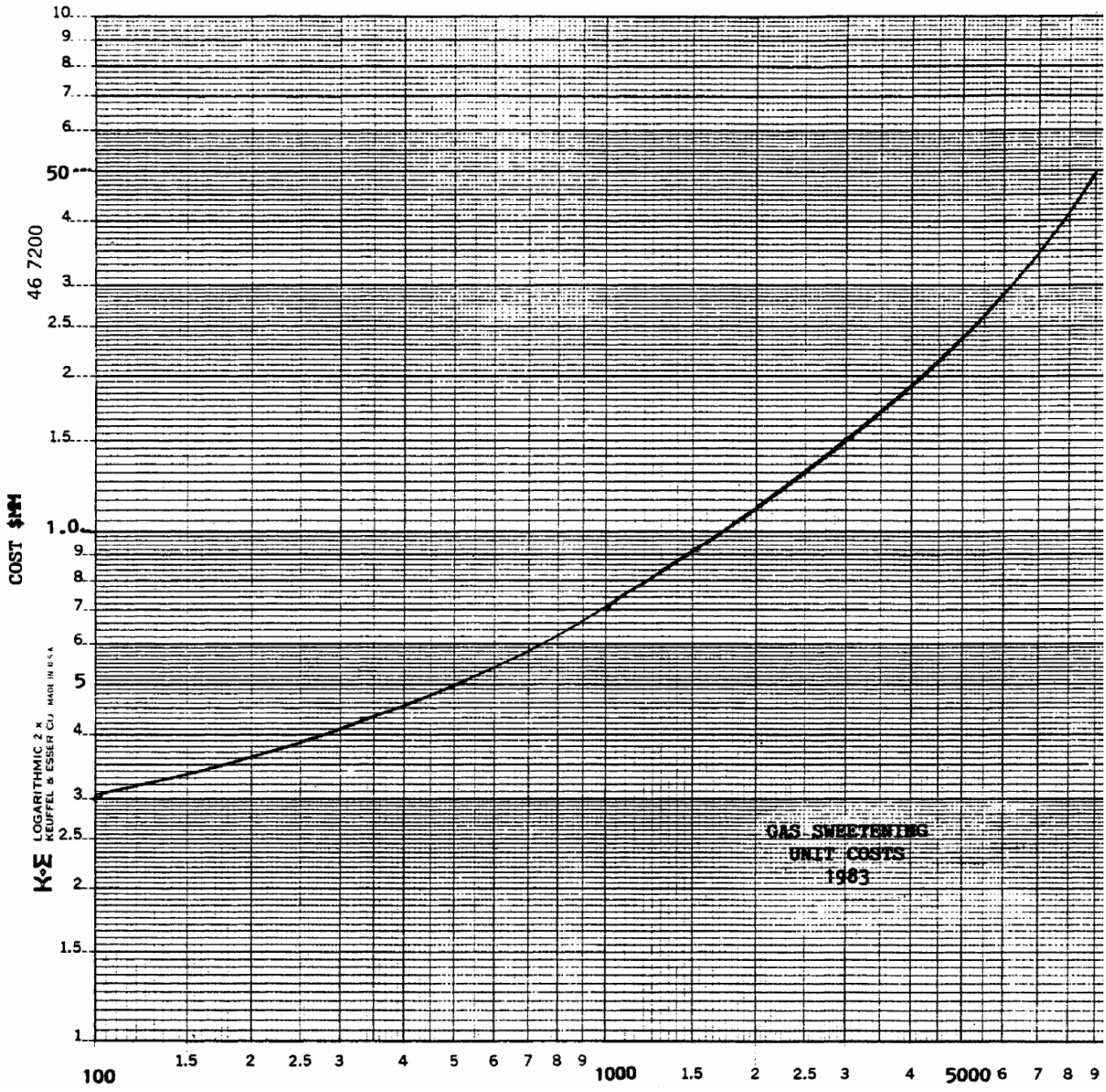


Figure 11.3: Gas Sweetening Unit Costs

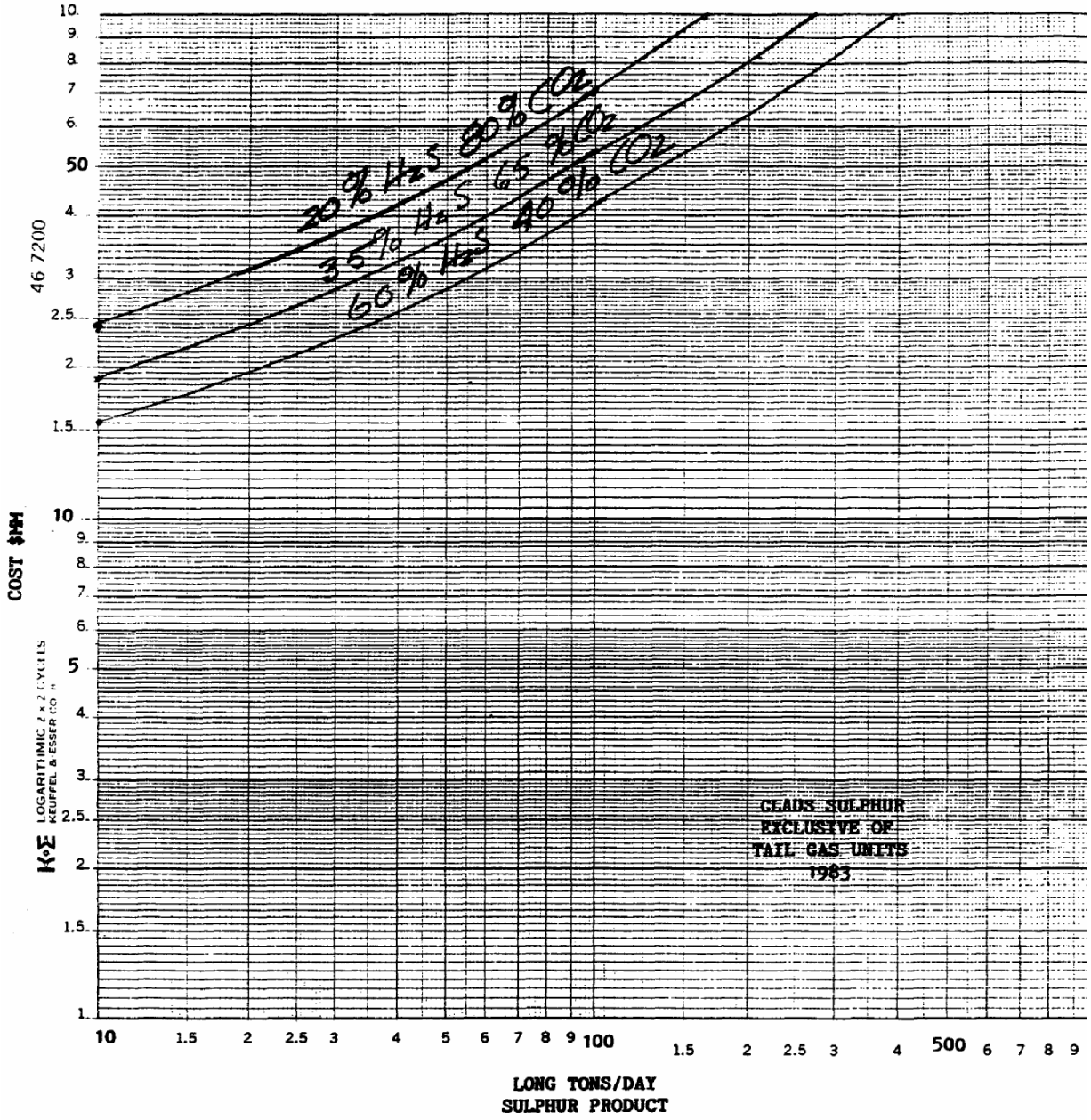


Figure 11.4: Claus Sulphur Exclusive of Tail Gas Units

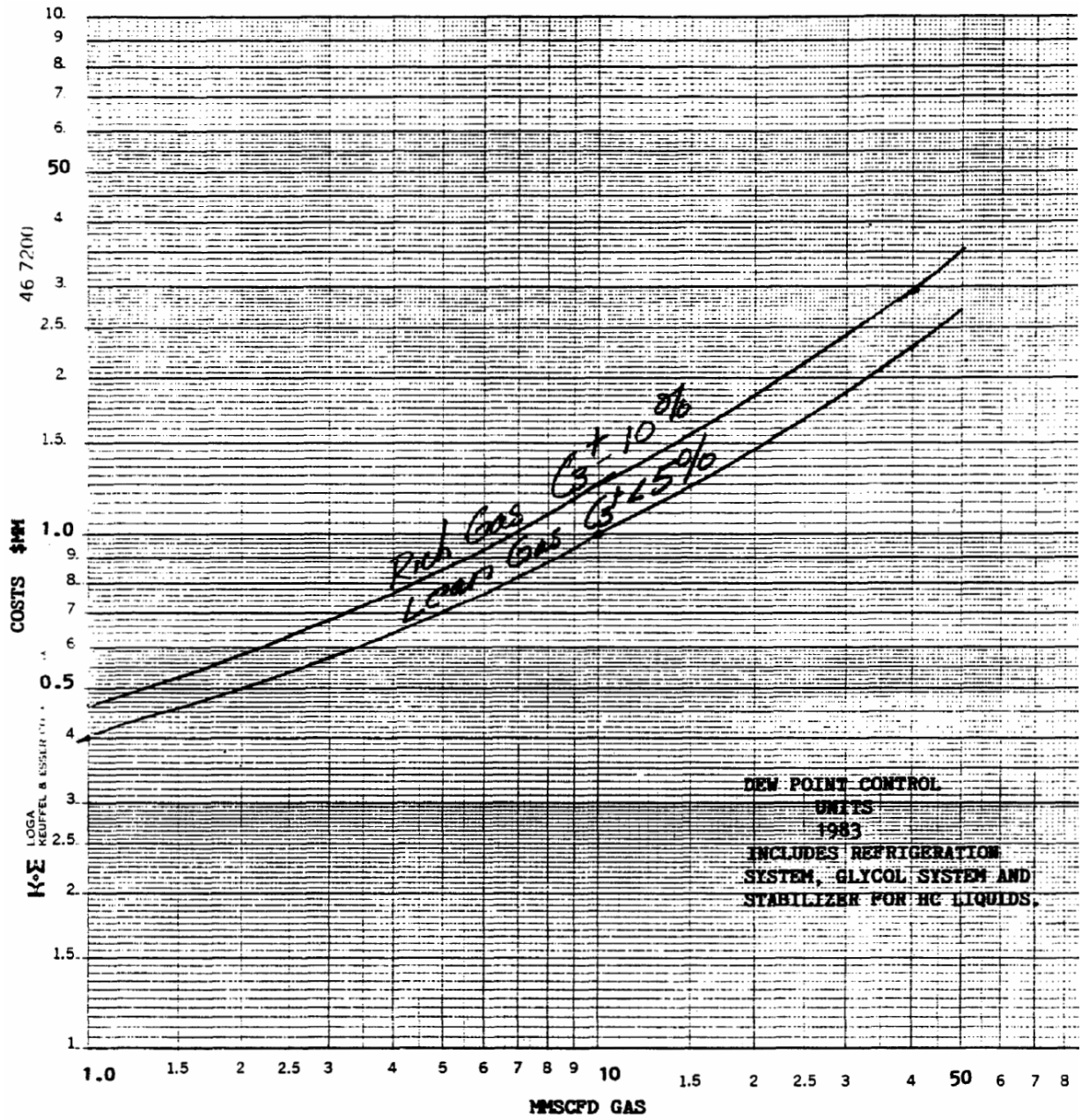


Figure 11.5: Dew Point Control Units

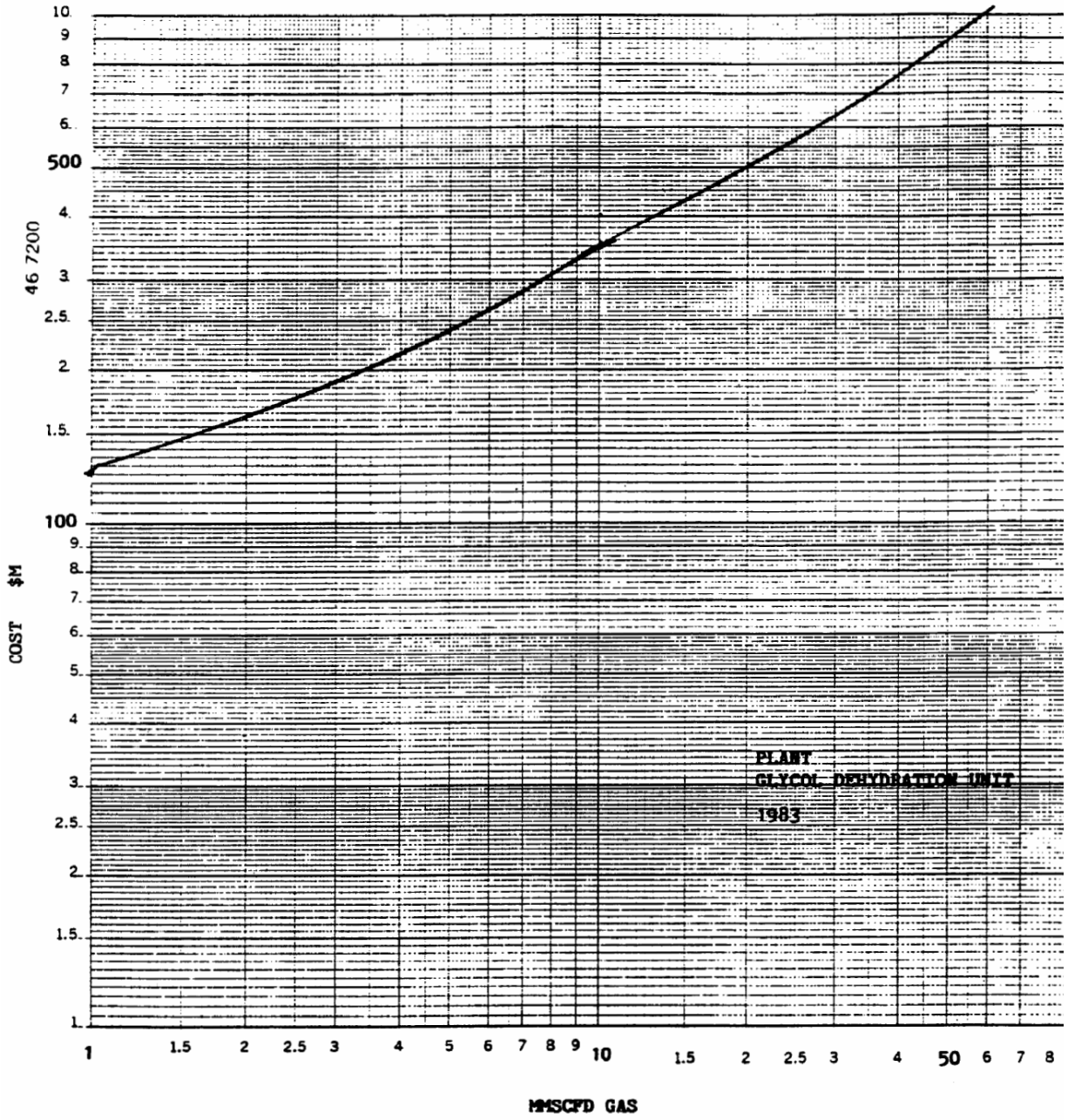


Figure 11.6: Plant Glycol Dehydration Unit

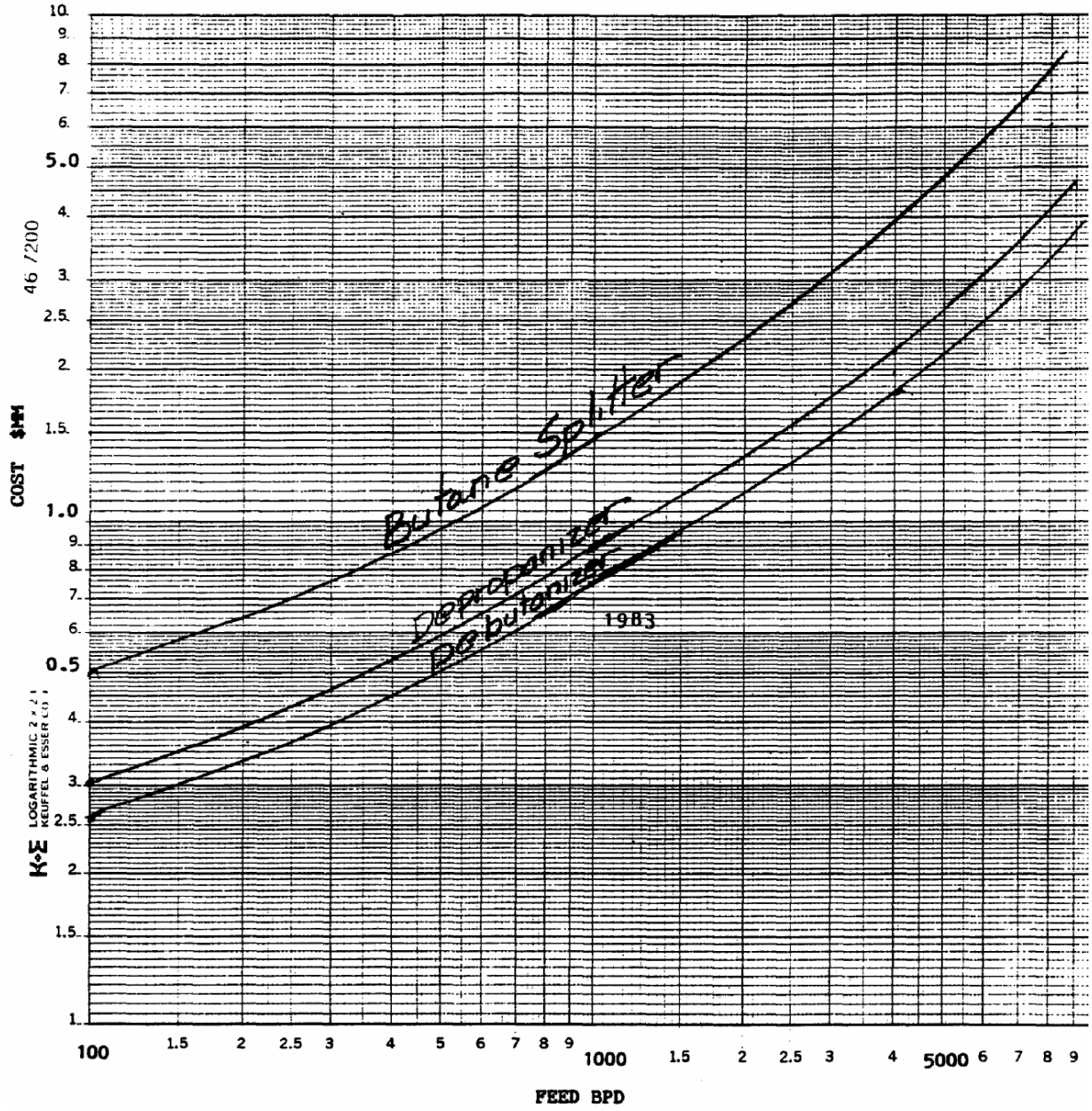


Figure 11.7: Fractionation Vs Feed BPD

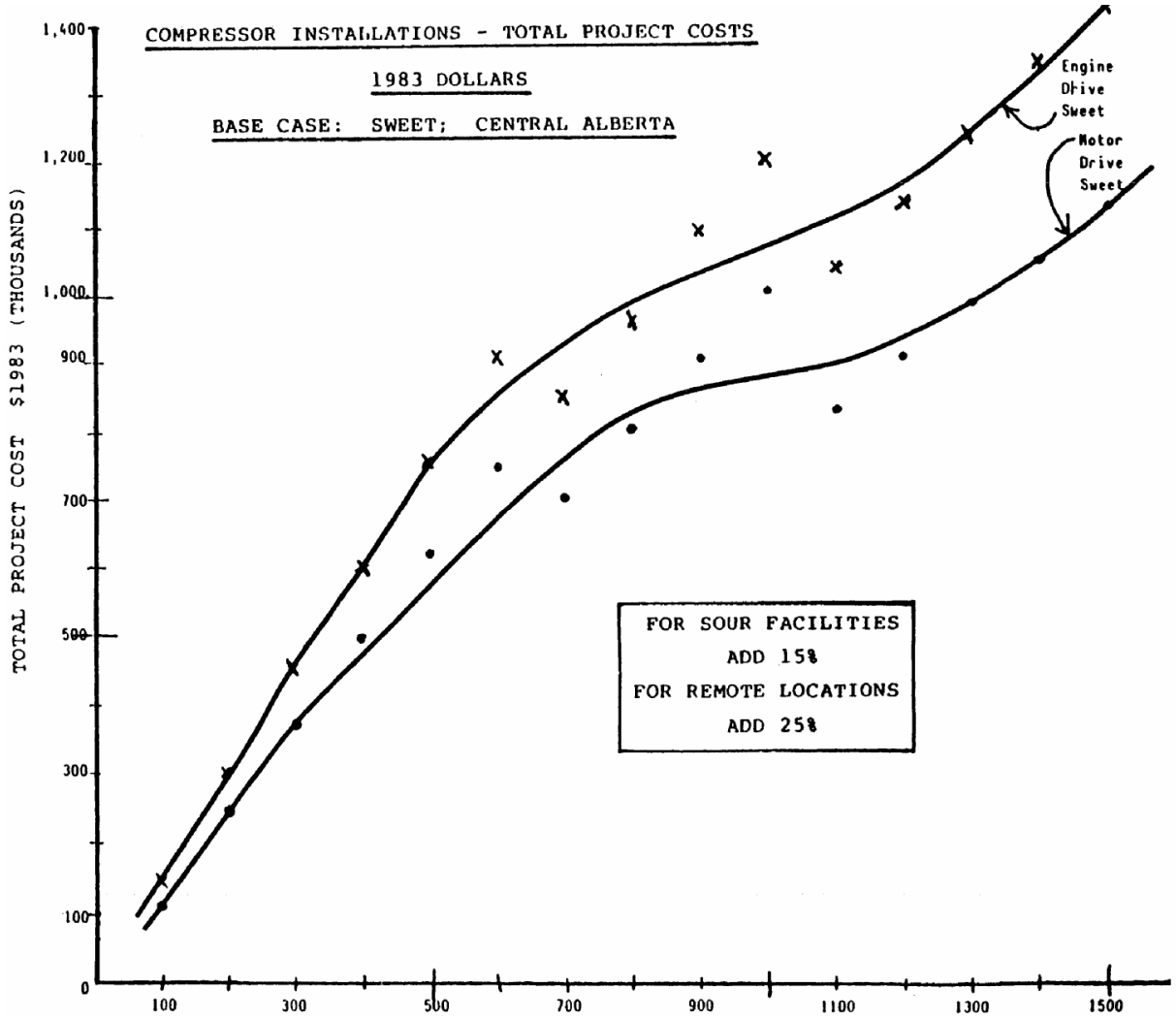


Figure 11.8: Compressor Installations – Total Project Costs

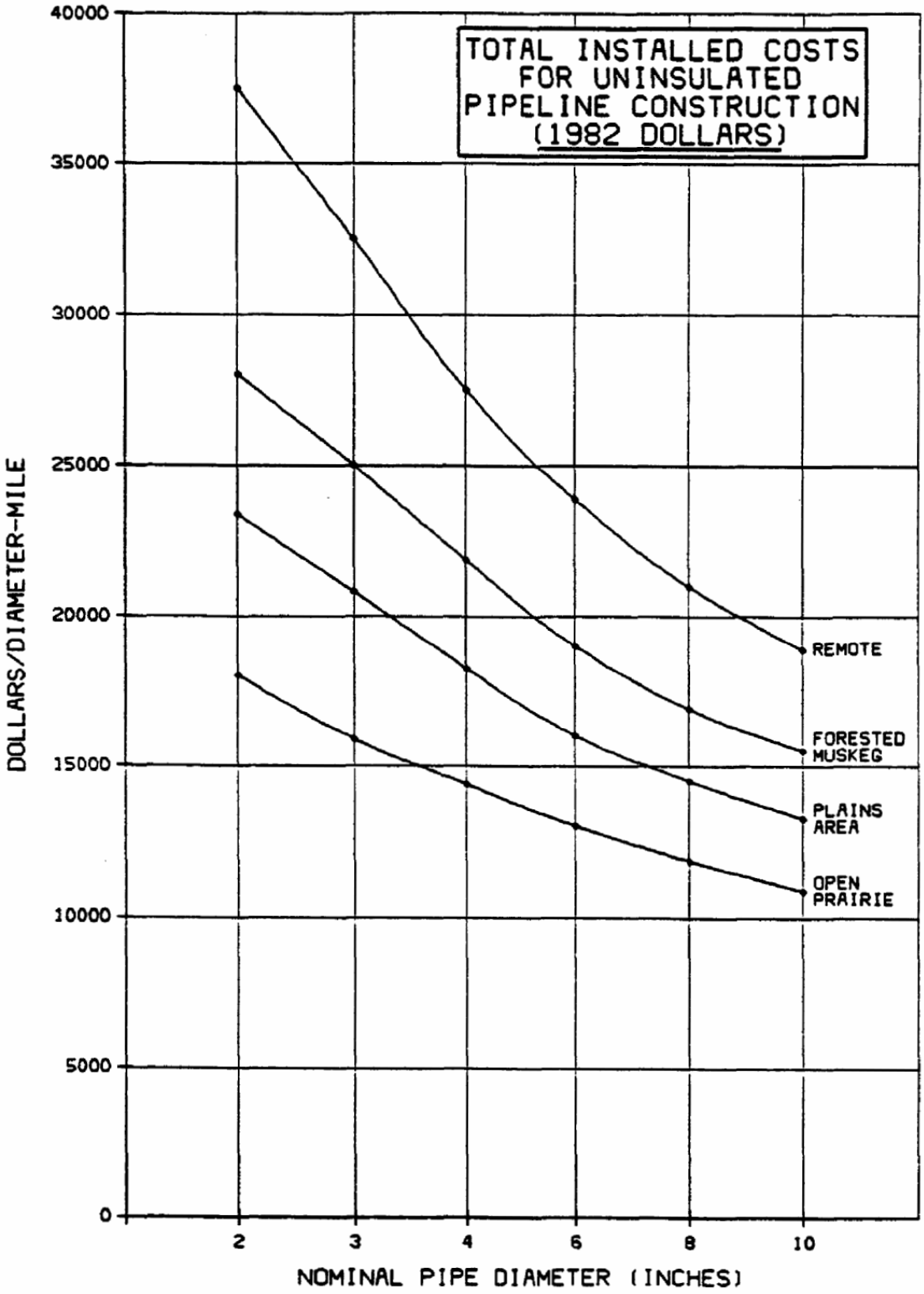


Figure 11.9: Total Installed Costs for Uninsulated Pipeline Construction

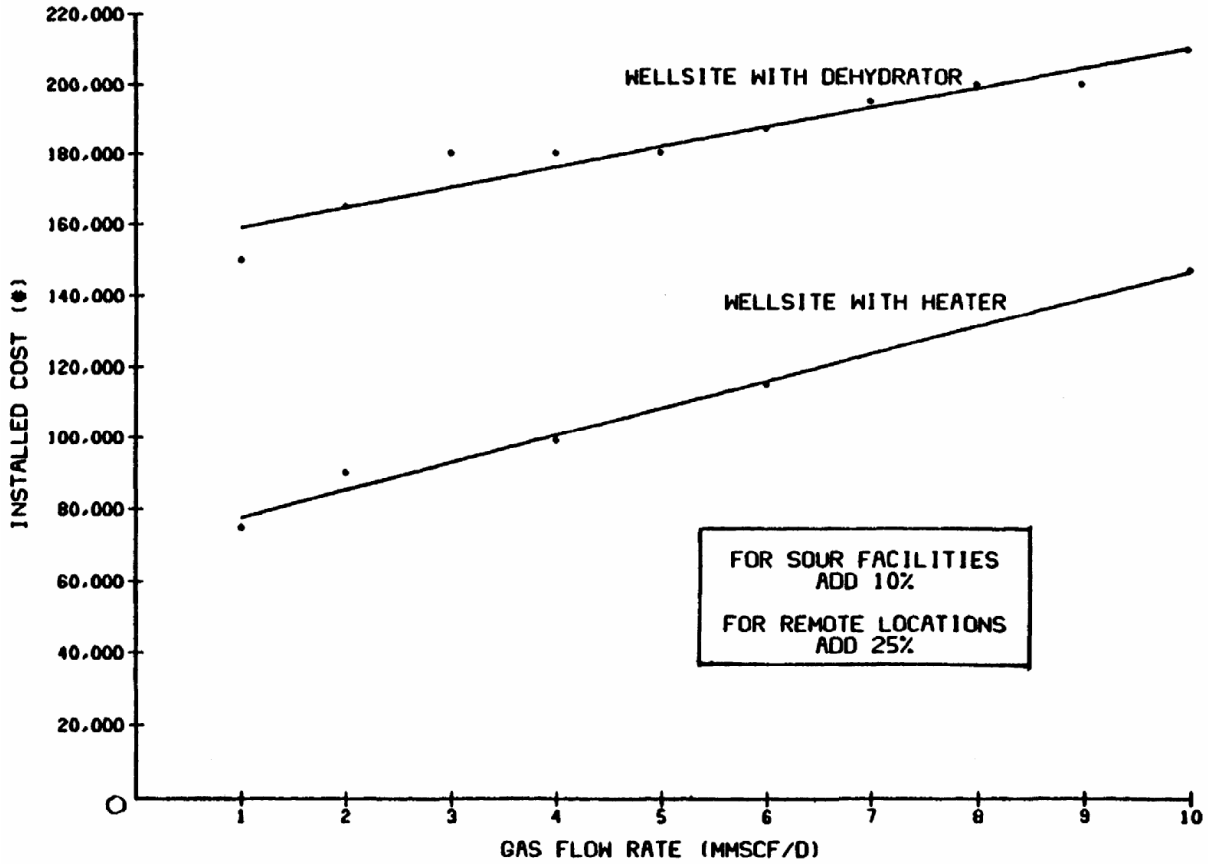


Figure 11.10: Wellsite Costs

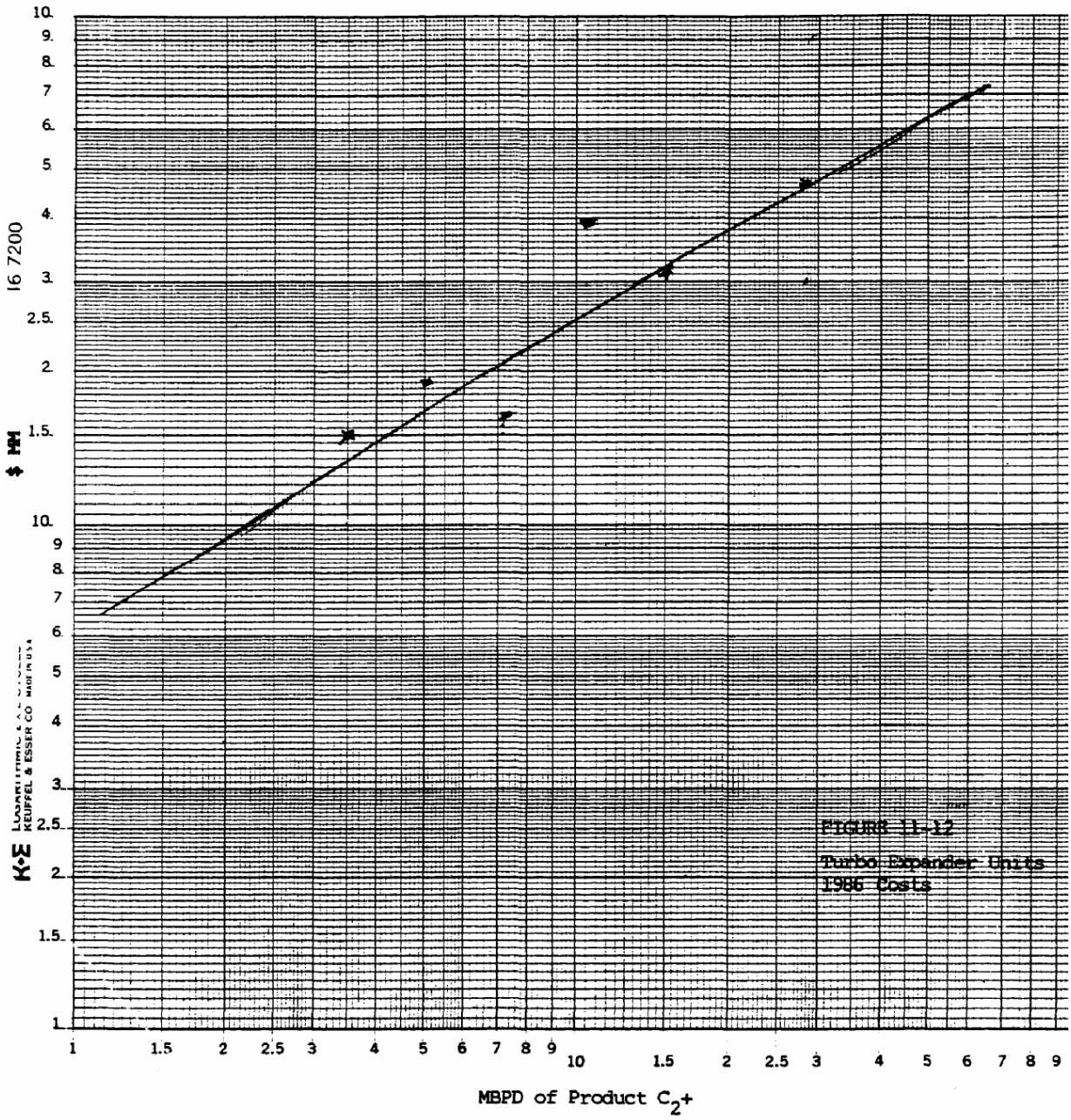


Figure 11.11: Turbo Expander Units (Costs)

The data on the Nelson Cost Index (now called the Nelson-Farrar Index) is given below. Please note that the design and productivity index does not seem to apply to gas plants, only the inflation index.

Another good index is the Chemical Engineering Plant Index as published by Chemical Engineering each month. The 1973 index was 140.8 and in February, 1984 it was 321. It is based on 1957-1959 being 100.

In 1983 the cost would be

$$\$10{,}000{,}000 = \frac{(1046.3)}{257} = \$40{,}700{,}000$$

The data for the Nelson Cost Index is shown as Fig. 11-10.

As mentioned above, the Nelson Index is primarily for plants. For pipelines a different index is used. It is the Morgan Index and the data for it is given in the following table.

Table 11.2: OGJ Morgan Oil-Pipeline Cost Indexes

	Yearly Indexes					1985 **			
Cost Component	**1980**	**1981**	**1982**	**1983**	**1984**	**1st Qtr**	**2nd Qtr**	**3rd Qtr**	**4th Qtr**
1. Steel line pipe	197	219	215	212	218	219	221	221	225
2. Line pipe fittings	193	214	209	207	213	214	215	215	219
3. Pipeline construction	174	195	207	214	219	220	221	222	224
4. Pipe coating	163	181	190	194	202	203	204	207	209
5. Masonry buildings	149	162	168	174	179	180	181	182	182
6. Metal or asbestos clad steel frame buildings	147	159	164	169	172	173	174	174	176
7. Stationary engines and reciprocating pumps	197	220	238	240	245	246	248	249	250
8. Electric motors - 100 hp & up and large centrifugal pumps	234	256	273	278	284	286	288	289	290
9. Portable and miscellaneous units	194	213	227	231	233	234	235	236	237
10. Electric starters and switchgear	156	172	183	190	195	196	197	198	199
11. Shop machinery and tools	192	215	234	246	256	258	260	262	264
12. Oil pipe and fittings in place	184	205	208	211	216	217	218	219	221
13. Steel storage tanks	190	208	218	220	226	227	228	229	231
14. Firewalls and miscellaneous	190	208	218	220	226	227	228	229	231
15. Radio and microwave equipment	143	157	165	171	175	176	177	178	179
16. Office furniture and equipment	150	162	172	180	185	186	187	188	189
17. Autos, tractors, and trucks	168	183	192	201	208	210	212	214	216
18. Other work equipment	193	213	230	242	252	253	254	255	257
Total Composite Index	**185**	**205**	**210**	**212**	**216**	**219**	**220**	**221**	**224**

** This index seems to have been discontinued; however, recent pipeline construction cost can be obtained from the article 'Pipeline Economics", Oil & Gas Journal, November 28, 1988, P.33. This gives current costs and a 10 year history of same.

Another record of the change of costs with time is available from the Canadian Petroleum Association. The following table gives the data that have been developed by them. The year 1970 is taken as 100.

Another index that is used is the "Canadian Chemical and Petrochemical Plant".

Table 11.3: Alberta Oil and Gas Industry Cost Escalation Indices

Cost Category	1971	1972	1973	1974	1975	1976	1977	1978	1979	1980	1981	1982
Geological and Geophysical (Field)	107	113	131	134	166	186	205	238	258	282	302	322
Exploration Drilling and Development Drilling	103	115	132	157	185	211	232	271	326	405	448	467
Field Equip., Secondary Rec. & Press. Maint.	107	119	142	180	213	240	269	290	325	357	402	436
Natural Gas Plant Construction	113	125	139	159	188	261	335	408	457	503	568	620
Total Capital Costs	108	119	136	163	193	227	251	285	331	392	439	473
Well Operating Costs†	113	127	157	165	207	245	293	334	386	414	434	461
Gas plant Operating Costs‡	119	127	129	152	193	239	284	338	361	436	492	514
Total Operating Costs	**115**	**127**	**147**	**160**	**202**	**243**	**290**	**335**	**379**	**420**	**448**	**473**

† Costs per well, including flow lines and related facilities.

‡ Costs per unit volume of marketable gas production from plants.

Table 11.4: Nelson-Farrar Cost Indexes

Refinery Construction (1946 Basis)								
	1962	**1973**	**1986**	**1987**	**1988**	**Oct. 1988**	**Sep. 1989**	**Oct. 1989**
Pumps, compressors, etc.	222.5	346.9	982.6	995.0	1027.7	1043.3	1095.2	1100.0
Electrical machinery	189.5	220.2	504.6	511.3	519.8	521.8	535.0	536.4
Internal-comb. engines	183.4	238.3	676.9	689.4	702.3	707.9	731.2	733.7
Instruments	214.8	338.0	733.3	744.2	748.8	750.1	778.2	781.9
Heat exchangers	183.6	313.7	524.7	575.2	664.1	707.6	744.1	744.1
Misc. equip. average	198.8	291.4	684.4	703.1	732.5	746.1	776.8	779.2
Materials component	205.9	-292.3	730.0	748.9	802.8	816.4	827.9	827.4
Labor component	258.8	585.2	1,330.0	1370.0	1405.6	1420.5	1443.5	1458.6
Nelson Refinery (Inflation) Index	237.6	468.0	1,089.9	1121.5	1164.5	1178.8	1197.2	1206.2
Refinery Operating (1956 Basis)								
Fuel cost	100.9	186.8	548.8	554.1	445.2	384.0	485.7	518.4
Labor cost	93.9	102.9	259.4	253.5	260.8	275.2	247.3	257.1
Wages	123.9	221.2	651.8	663.2	689.8	711.1	709.3	707.7
Productivity	131.8	214.9	253.8	261.8	264.6	258.4	286.8	275.3
Invest., maint., etc.	121.7	193.0	460.0	473.2	489.3	495.3	503.0	506.8
Chemical costs	96.7	117.3	206.7	192.8	213.9	224.5	226.2	225.8
Nelson operating indexes Refinery	103.7	125.7	369.0	371.8	373.7	377.3	379.4	387.7
Process units **	103.6	168.0	430.8	435.6	405.3	390.3	420.3	435.9

** Add separate index(es) for chemicals, if any are used. See current Quarterly Costimating, first issue, months of January, April, July. and October.

Note: These indexes are published in the first issue of each month. They are compiled by Gerald L. Farrar. Journal Contributing Editor.

Indexes of selected individual items of equipment and materials are also published on the Costimating page in the first issue of the months of January. April. July. and October.

One process unit not given in the figures is inlet separation. For this, a rough cost of about $400/barrel of capacity (in 1982) can be used. When all process units have been costed add 25%-30% for offsite ie. office, shop, warehouse, utilities etc.

As well as the process units, the well site facilities and the gathering pipelines, several other items need to be calculated. These include:

a) Development wells - The cost of these should be obtained from the drilling department.

b) Roads - The cost of roads for year round access to all wellsites and other field facilities should be included. Proper recognition of the amount of road work already built to drill the well should be made. The cost of an all weather road, of course, varies a great deal depending on the terrain, gravel availability, etc. The normal range is from $10,000/mile or less in the plains area to up to $30,000/mile in areas of hills, muskeg, etc.

c) Catalyst chemicals and initial charges of certain raw materials - Note that in some cases this cost can be very large; for example, a crude pipeline has to be filled and this fill is required for the length of the project. The costs for these items are determined by take-off generally.

d) Initial spare parts, supplies, etc. - If no data is available, about 1% of plant investment should be used.

e) Equipment supplied directly by the owner - such as maintenance equipment and tools, office equipment, office furniture and supplies, safety supplies, communications equipment, laboratory equipment and supplies, etc. If costs for each item are not available, use about 1.5% of capital cost for the plant for this.

f) The owner's costs for engineering - special facilities, and for inspecting the contractor's work, both in the engineering office and in the field. This will vary from about 5% of capital cost for small jobs to about 1% for large jobs.

g) The owner's cost for managing the project - should be determined from the agreement for constructing the facility, if signed. If it is not, and the anticipated rate has not been determined, then use 1% of capital costs.

h) Startup costs - which include the cost of training all personnel to operate the plant. The total personnel for the plant at three months salary is probably a good minimum for this; often this charge is expensed, if possible.

i) The cost of purchasing the land should be included.

j) Interest during construction - Often the cost of borrowing money for the building of the facilities is added to the capital cost. This is taken as the cost of interest for obtaining the funds for the project for half the time it takes to build the project.

k) Working capital - Some of this is capitalized. A figure of 3 months operating cost is often used.

Another item to consider is the location of the work and the productivity expected from the construction labour. The data given in the curves are for Western Canada, in the Prairie region. If the facility is in a different area, the productivity factors should be allowed for in the plant and pipeline work. About 30 - 35% of the plant cost and 45 - 50% of the pipeline and wellsite cost are labour.

One final note is that costs for an existing unit are very good source of cost data even if the size is different, Usually the 0.6 rule can be used for size ie.

Equation 11-2

$$New\ Cost = Cost\ of\ \ Existing\ Unit\left(\frac{size\ of\ \ new\ unit}{size\ of\ \ old\ unit}\right)^{0.6}$$

This 0.6 should be increased 0.7 or 0.8 if multiple trains have to be put in.

11.4.3 Appropriation or Budget Estimate

An improvement in accuracy can be obtained if the actual preliminary process design is done so that the major equipment is sized and used for calculating the cost. This is done by using the cost of each piece of major equipment as indicated in the method described below.

In this estimate all major pieces of equipment are sized and then priced, either by quotation or from available data on cost for the particular equipment, such as j/lb for pressure vessels, $/HP for compressors, $/ft2 for exchangers, etc. The cost of each of these pieces of equipment is summed up and the resulting number multiplied by a number that generally varies from 2.5 to 4.0 depending on the size and location of the facility. The table below gives the range to use.

Plant Cost/$MM	Multiplying Factor Δ
less than 10	3.5
10—25	3.2
25+	3.0

When in an existing plant and close by the existing units, add up to 0.5 to the multiplying factor. The accuracy of an estimate done this way is probably -10% to +20%. A variation on this type of estimate, together with some excellent cost data, is given in the article by K. M. Guthrie.

Another variation of the major equipment method is to find the total installed cost for each major piece of equipment by multiplying the cost of each piece of equipment by the factor for its installation as given in Table 11.1, then summing these costs and adding 25% to 30% for direct field costs to find the total cost.

Table 11.5: Installation Factors for Major Equipment

Exchangers	2.0
Compressors	2.5
Expanders	2.5
Pumps	3.0
Vessels	2.0
Towers	2.3
Field Fab Tanks	2.0

Miscellaneous Equipment	2.5

The advantage of this method over the Preliminary Budget Estimate method is that the major engineering has been done and special engineering innovations can be taken into account.

11.4.4 The Definitive or Detailed Estimate

This is an estimate that is used for firm price bidding on contracts or for cost control on projects if the contract has already been awarded because of time or other considerations. The engineering is usually well along, (about 40 - 50% done) quotes for all major pieces of equipment have been received, depending on the contract arrangements, most .of it purchased. The required amount of labour for the work is estimated and all the possible factors that can be estimated are included, such as productivity estimates, requirements for scheduled overtime, the need for construction camps, and all similar data. Also included in the detailed estimate are take-offs and the pricing of most of the material for the job such as pipe, electrical material, concrete, structural steel, etc. The requirements for such an estimate and the cost of doing it are given below.

The 'Detailed Estimate' is a takeoff type estimate that is done after the 'Design Specifications' have been issued and 40% to 80% of the detailed design engineering has been completed (The term Detailed Estimate is synonymous with Definitive Estimate.) The time and cost of producing a Detailed Estimate will be approximately ten times that of producing a 'Budget or Appropriation Estimate', and some thought should therefore be given to decide whether, in fact, the Detailed Estimate is necessary.

The usual reason for having a detailed Estimate prepared is to have a basis for comparison for cost control purposes during construction. This is usually done for reimbursable cost type contracts. (Lump Sum or Unit Price type contracts are capable of being controlled without a Detailed Estimate, as such an estimate has been done by the contractor to bid the job). Detailed Estimates are also sometimes required as a final check on costs before large sums are committed for construction, when trends indicate the need for a fresh look at the total project cost.

In large complex projects areas can be ready for Detailed Estimating at different times in the project. Each should be estimated as it reaches the appropriate time, but it should be remembered that the individual area estimates may be altered by overall project conditions (i.e. labour productivity) that are not known at the time the area estimates are prepared.

Whenever possible, the EPC Contractor's code of accounts and estimate format should be used for the Detailed Estimate in the Contractor's cost reporting system. Using the Contractor's code and format is the most efficient approach since the Contractor is most familiar with his own procedures.

The Detailed Estimate is a detailed, step by step prediction of the way the job will be executed, expressed in terms of quantities and costs. It is intended to reflect average performance on the part of the Contractor.

Therefore, it provides a meaningful yardstick with which to measure the contractor's performance. The Detailed Estimate normally forms the basis of the cost portion of a unilateral incentive plan.

The timing of the release of the estimate details to the Contractor, by the Contractor's estimating department, is critical. Proper cost control dictates that the control budget for any component be available before that component is committed. Often this can only be accomplished by piecemeal release of the estimate details as they are completed, on a priority schedule basis. The contractor normally bases his estimate on preliminary quotations and layouts and preliminary execution plans.

The information necessary in order to produce a Detailed Estimate is essentially the detailed design information that will become available as a result of the detailed design engineering being completed. These information requirements are detailed as follows:

a) Changes (if any) to the information on which the Appropriation Estimate was based.

- Project Scope.
- Site Data.
- Process Description (Final P & ID's).
- Plot Plans (Facilities and Units).
- Project Master Schedule (including risk analysis).
- Basic Design Documents/Specifications approved for design.

The Civil and Structural disciplines will generally be more advanced than the Mechanical and Electrical disciplines and information flow necessary to develop the Detailed Estimate will therefore follow this sequence, rather than in the order listed below.

6. Equipment

- Mechanical flow sheets.
- Completed specifications and data sheets on all items of major .equipment
- Firm bids/quotes received (price and delivery).
- Completed unit layout and arrangement drawings.
- Auxiliary equipment specified and priced.

7. Piping

- Detailed P & ID's completed.
- Completed piping layouts, isometrics, spool sheets, line lists, and bills of material.
- Scale model (if used).

8. Concrete and Civil

- Detailed drawings of foundations, concrete structures, and civil works completed.
- Material takeoff quantities defined.
- Prices of borrow material from established quarries.
- Proving results of local sources of sand and stone.

9. Structural Steel

- Detailed drawings of process buildings, pipe supports, operating platforms, stairways, and ladders.
- Changes called for by Operation & Maintenance Groups are finalized.
- Model (if used) reviewed and approved.

10. Instrumentation

- Detailed instrumentation drawings (auxiliary air supply, tubing valves, spool pieces, etc.
- Instrumentation lists, quantities and prices.

11. Electrical
    - Completed specifications and data sheets on all items of electrical equipment.
    - Firm bids/quotes received (price and delivery).
    - Detailed construction drawings.
12. Insulation and Paint
    - Final specification of insulation and paint.
    - List of insulated items.
    - Detailed drawings of insulated items.
13. Buildings and Structures
    - Detailed structural drawings for buildings.
    - Architectural details for buildings.
    - Bills of material.
14. Offsite Facilities
    - Final utility systems. Detailed layouts and equipment drawings. Utility contracts signed.
    - Final auxiliary systems. Detailed drawings and layouts for townsite, camps, roads, airfields, etc.
    - Final storage and tankage requirements. Detailed drawings.
    - Detailed drawings for offsite piping layouts.

In addition to the detailed drawings of the facility itself, the following studies should be completed:

15. Complete Survey of Economic Conditions and Market and Labour Data at Site
    - Labour productivity analysis.
    - Labour wage rates over extended project life.
    - Competing projects likely to affect labour market.
    - Material costs in the area.
    - Vendor shop loadings.
    - Labour supply and local content requirements.
    - Supervisory availability.
    - Escalation projections for wages, material and equipment.
16. Home Office Costs
    - Staffing requirements by disciplines, man-hours and rates (including burdens).
    - Home office indirects.
17. Field Administration
    - Staffing requirements (man-hours, rates and burdens).
    - Indirects.

18. Field Indirects
    - Detailed specification of temporary construction facilities, requirements and costs.
    - Construction plant and equipment.
    - Definition of construction logistics.
19. Construction Contracts
    - Construction schedule (Level II).
    - Manpower loading.
    - Construction approach and plan.
    - Major contractual packages.
    - Type of Contract(s) to be used (RC, LS, UP).
20. Sub-Contracts
    - Estimates based on unit quantities, unit rates, or firm quotes.
21. Owner's Costs
    - Taxes.
    - Duties and drawbacks.
    - Interest.
    - Insurance.
    - Exchange rates.
22. Spare Parts Lists Completed
23. All Major Permit Approvals Obtained
24. Commissioning Plan

The estimating technique used for Detailed Estimating is essentially a "take-off" approach of materials, equipment and labour units.

This approach is fundamentally different from the factored techniques that are used for earlier estimates. In the take-off method, the estimate is prepared "from the bottom up" rather than "from the top down", as is the case with factored estimating. This is the reason why the take-off approach is very time consuming and costly. Below follows a summary of the method.

Home Office - detailed man-hour estimates by disciplines.

Equipment - purchase orders placed or quotations available.

Commodities - detailed material take-offs with quoted unit costs.

Erection - standard unit construction man-hours for equipment and commodities costed at rates estimated for the project.

Indirect Construction - detailed estimates for temporary construction facilities, detailed listing of construction equipment requirements costed at quoted rates.

Construction - detailed supervisory staffing requirements.

Contingency & Escalation - detailed analysis by each account.

Special Factors - allowances for special factors and conditions could affect costs.

The time for doing this work is 6 - 8 months after awarding the contract or 4 - 6 months after a bid request is sent out, depending on the detail in the bid request. The accuracy depends on the amount of items that are fixed in price, probably 7% to 15% or better.

11.4.5 Accuracy of Estimate

There is now a lot of interest in doing a probability analysis of the accuracy of estimates, particularly by financial people. This can be done by various statistical analysis techniques. This requires the review of each part of an estimate and putting a maximum and minimum increase or decrease on the cost of that part of the estimate, then using a known statistical procedure to develop the most probable cost and then the probability that the cost will be below or above the final estimated cost. A more detailed discussion on this subject is given in the next few paragraphs.

There are many uncertainties connected with an estimate, however, the total expected cost can be separated into three components.

11.4.5.1 Known Cost

Known cost is the base estimate calculated from the Basis of Estimate documents. Also included are material growth, take-off allowance, unbudgeted bulk material allowance, construction spares, project completion costs and pre-commissioning activity costs.

11.4.5.2 Known Uncertainty

Provision for Known Uncertainty is usually included in the contingency. The contingency analysis system for known uncertainties uses the RISKAN II program (a proprietary software package licensed by Control Data Corporation) as a basis. RISKAN II is designed to provide contingency analysis by using Monte Carlo simulation and sensitivity analysis techniques. Additional in-house input and output analyses and cost profile graphics are used to provide comprehensive contingency analysis.

System input includes estimate data used to produce the Overall Estimate Summary in conjunction with accuracy statements (risk memoranda) obtained from each engineering discipline, project management, process, procurement, construction and cost Services.

The overall project is first separated into significant estimate input elements. The combination of these elements into the total project cost is then defined by a mathematical model. An appropriate range and distribution of element values is specified based on the definition, take-off and pricing from the risk memoranda.

The Monte Carlo simulation uses the mathematical model to recreate the estimated cost of the project. Values for each estimate element are selected at random from the user specified range and distribution. The program then calculates the resulting project cost a significant number of times. These costs are analyzed and the results are summarized.

This method of contingency analysis does not consider the effect of unknown uncertainty. Unknown uncertainty must be evaluated as a separate component of the final contingency.

11.4.5.3 Unknown Uncertainty

Allowance for unknown uncertainty includes those pervasive costs which are inappropriate for RISKAN II analysis. This allowance is determined by conventional analytical methods. Risk items that are typically included in this category and excluded from the RISKAN II analysis are listed below:

Economic Changes

- escalation/inflation.
- currency fluctuations.
- altered market conditions.

Political Changes

- unknown results of a significant political change.

Actions by Outside Agencies

- technical (changes in Codes, Regulations, etc.).
- commercial (changes in import duty rates, tax rates and/or application, etc.).
- labor (extended work stoppage at critical vendor, preassembly yard or site).

Non-performance by Vendor or Sub-contractor

Changes by Client

- in specifications.
- in scope.
- in coordination procedures, etc.

Site

- significant geotechnical surprise uncovered by detailed marine/soil investigation.

Force Majeure Events

11.4.6 Final Estimated Contingency Calculation

Using the above data, here is an example of calculating the recommended contingency.

25. Known Uncertainty

Based on the following Operations Cost Profile, a 20% overrun probability is recommended which indicates a contingency for known uncertainty of 10.9%.

	Base Cost % **	Recommended Contingency %
Equipment	10.5	5.4
Materials	12.2	13.5
Subcontracts	27.2	5.8
Construction	36.4	14.6
Home Office & Fee	7.9	16.6
Miscellaneous	5.8	7.6
Total	**100.0**	**10.9**

** The range for items making these costs up on typical jobs is given in Figure 11.12.

26. Unknown Uncertainty

	%
Schedule Slippage (3 months)	1.0
Strike	1.2

Labour Shortage	0.2
Currency Exchange Risk	0.6
Regulatory Risk	0.5
Miscellaneous	0.6
Total	**4.1**

27. Total Recommended Contingency

Total Recommended Contingency = 15.0

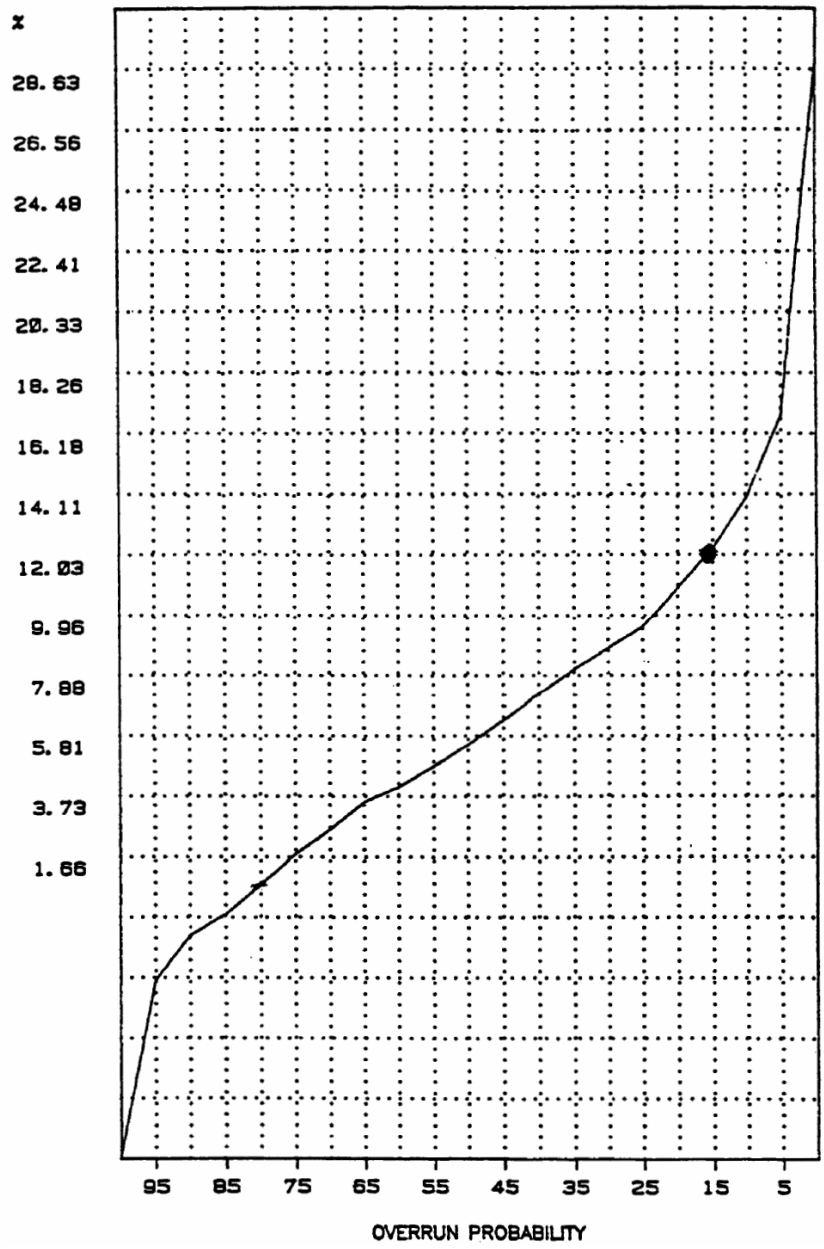


Figure 11.12: Operations Cost Profile

11.5 Operating Costs

A very rough estimate for operating cost is 6-10% of the capital cost for the annual operating cost, the 6% being for a simple sweet gas plant and the 10% for a complex sour gas plant exclusive of raw material or feed cost.

Another way is as follows:

divide the cost into raw material costs, fixed costs, variable costs and special costs.

b) Raw Materials - In gas processing facilities the raw materials are normally not assigned a cost. However, in pipeline straddle plants the raw material is natural gas, which is purchased and thus does have a cost.

c) Fixed Costs - There are costs that do not vary with throughput such as supervisory, technical and clerical staff, operating labour, maintenance, taxes and insurance, etc.

- Supervisory, Technical and Clerical Staff - A plant operating organization should be developed. The minimum requirement is probably three (plant superintendent, process foreman and clerk) for anything except a very small plant. Larger plants require several more personnel. The number of staff is multiplied by $50,000/yr (1984) for the total cost. This cost includes all benefits, etc.
- Operating Personnel - An attempt should be made to determine the area that a single operator can handle. Generally, this is set by the number of pieces of equipment that need to be checked. Several different process units require more men than a plant of the same cost but having a single process unit. After the number to cover the plant has been determined, quantity is multiplied by 4 or 4.2 to give 24 hour coverage. However, the number of different process units is probably the best guide. The number required is multiplied by $45,000 (1984) to find the total cost.

Table 11.6: Typical Chemical Plant Costs

Cost of Components of Typical Chemical Plant		
Category	**Percent of direct total**	**Range, %**
Foundations	4	1-8
Equipment	39	26-46
Insulation	4	1 -5
Site work	4	1-9
Buildings	10	6-20
Duct work	1	0-3
PWR wiring	8	7-12
Instruments	8	5-10
Piping	22	12.29
Major Cost Categories in a Typical Chemical Plant		
Category	**Percent of Project Range**	
Engineering design	7-13	
Engineering control[¥]	1-3	
Indirect field costs	10-18	
Field labor	25-35	

Field material	15-20
Fabricated equipment	25-35

[¥] Engineering control includes the following:
Construction home office expense and drawing reproduction.
Quality assurance engineering and vendor inspection.
Cost control and estimating.

- Maintenance - This is usually estimated on a fixed percentage of capital investment. Hydrocarbon Processing often publishes data for the costs.

  For sour gas units use about 3-4% of capital costs.

  For sweet gas units use 2-3% of capital costs.

  This includes supervisory, maintenance staff, maintenance labour and maintenance material.

- Property Taxes and Insurance - The cost for this runs about 1.5% of investment (This covers only municipal and similar type taxes, not income taxes).

- Miscellaneous Cost - This covers telephone calls, stationery, office operation, vehicles, etc. Use 0.25 to 0.5% of capital cost/year.

d) Variable Costs - These are costs that vary with the throughput of the plant.

- Chemicals - The consumption and costs of these are shown in Table 11.7

Table 11.7: Chemical Consumption and Costs

	Consumption	$/kg [**]
Sweetening Chemicals		
Monoethanol Amine	1 lb/% of acid gas/MMSCF	1.87
Diethanol Amine	0.5 lb/% of acid gas/MMSCF	1.89
Dissoproply Amine (98% purity)		2.025
Sulpholane (3% Water)	3.0 lb/MMSCF	5.85
Methyl Diethanol Amine	Like DEA	4.35
Dew Point Chemicals		
[§]Monoethylene Glycol	1 lb/MMSCF	1.35
[†]Diethylene Glycol	4-6 lb/MMSCF	1.42
[†]Triethylene Glycol		2.09
Calcium Chloride	2 lbs/MMSCF	1.13
Propane	1 lb/MMSCF	0.22/Imperial Gallon
Other Common Chemicals		
Activated Alumina (sulphur catalyst)		1.23
Silica Gel (desiccant service)		3.86
Molecular Sieve 13x		4.75
Molecular Sieve 4A		3.75
Methanol	non recoverable	1.10/Imperial Gallon

[**] These are January 1989 Alberta costs in Tank car quantities.

§ Dewpoint suppression service.

† Dehydration service sour gas lower for sweet gas.

Other chemicals, such as antifoams, boiler and water treating chemicals; estimate directly or assume they cost about 10% of major chemicals.

- Fuels and Utilities - Generally only electricity purchased. The cost of this should be obtained from the power company, but in Alberta it is usually low 2.5 cents to 3.0 cents per kwh/hr. Sometimes water is purchased from a city supply. The cost is about $0.20/1000 imperial gallons.

  Fuel gas is normally supplied internally and this is valued at sales gas value 25 years from plant startup and thus is taken at discount of sales gas value. Also, if there is a lot of by-product, high fuel gas consumption allows an increase in the recovery of these. This, of course, is not correct in straddle plants or other plants not processing gas or products directly from a field. In plants not on a field fuel costs can often run to 1/2 the operating cost or more.

e) Special costs

- Head Office cost - Often to operate a plant, a company applies a 10 to 12% fee on all operating costs exclusive of raw material costs to cover head office functions.
- Royalties - For all petroleum products that are considered to be obtained directly from a reservoir the various provincial and federal governments in the territories take a royalty. This varies from province to province and also depends whether it is "old" or new gas. It is necessary to check with governments involved to see what the royalty is. There is often a fee that is returned to the producer for processing royalty gas.
- BTU and Shrinkage Penalties - Sometimes if there is extensive recovery of LPG it is necessary to allow for the change in the BTU content of the sales gas and the change in the sales gas daily contract volume.

f) Depreciation or Capital Cost Allowance - This is only necessary for finding net income after taxes. The actual depreciation is added back to net income after taxes to obtain the cash flow. Current income tax depreciation rates are:

Wellsite facilities	25%
Gathering Pipelines	25%
Product Pipelines	6%
Processing Plants	25%
Development Wells	30%

g) Taxes - These are continually changing. It is necessary to check with tax people for rates and what is allowed in deductions. Recent tax rate (1990 are given below).

Federal Income Taxes	28% of Net Income before Royalty Deduction.
Provincial Income Taxes	15% of Net Income after Royalty Deduction.

There is also a 25% resource allowance deduction that also has to be taken into account.

Both (e) and (f) cost are continually changed by governments, thus for detailed economical calculation current data should be obtained.

11.6 Prices for Products

The prices for products produced are very important in the economic calculations. These, of course, vary greatly depending on the market situation, governmental policy, world crude prices, etc. Also, the prices in future years depend a great deal on assumed inflation and interest rates. The most recent estimates of these values should be used in any economic evaluation. Also, one or two different scenarios besides the base case should probably be investigated unless a very quick payout is to be realized.

Following are predicted values for some of the gas and gas products as of the end of 1988.

Table 11.8: Predicted Gas Product Prices

		Years						
Item	**Units**	**1989**	**1990**	**1991**	**1992**	**1993**	**1994**	**1995**
Natural Gas								
Alberta Average Market Price	$Cdn/MM BTU	1.70	1-90	2.35	2.75	3.15	3.80	4.45
LPG								
Alberta Ethane	$/Cdn/BBL	6.00	6.75	8.25	9.75	11.00	13.25	15.50
** Alberta Propane	$/Cdn/BBL	9.00	10.25	12.50	14.50	16.50	19.75	23.75
† Alberta Butane	$/Cdn/BBL	15.75	17.50	20.00	22.00	24.25	27.75	31.75
Sulphur at Plant								
Gate	$Cdn/ L.T.	78.0	81.5	85.0	89.5	94.0	98.5	103.5
Edmonton Light								
Crude Oil	$Cdn/BBL	21.00	23.50	26.50	29.50	32.50	37.00	42.00
Texas Crude	$US/BBL	18.25	20.00	22.75	25.00	27.50	31.50	36.00

** Normally 60 — 65% of Crude

† Normally 80 — 85% of Crude

11.7 Information Required to Evaluate a Project

Here is an outline of the information generally required to evaluate a project.

General

- project life.
- starting date.
- location – geographic.
- nature of the project - new business, cost saving, incremental.

Revenue Information

- volumes plus a forecast, high and low case.
- prices plus a forecast, high and low case.

- market forecast.
- market share forecast.
- timing, ie. date of first and last income streams.
- environmental factors which may affect the revenue stream, ie. market or political considerations.

Capital Costs

- magnitude.
- value at a specific point in time.
- description as to exactly what is being acquired (e.g. capital costs for a building for $10,000,000 may be $1,000,000 land, $7,000,000 buildings, $1,000,000 in project management fees, $1,000,000 in architectural fees).
- timing, ie. construction start and end dates.
- capital cost inflation factors.
- estimates of resale values.

Operating Costs

- units and a volume forecast.
- prices and a volume forecast.
- raw materials.
- labour.
- utilities.
- fixed costs.
- maintenance.
- supervision.
- timing information.

Project Alternatives

- lease vs. buy.
- trucking vs. pipeline.
- build a plant vs. pay a processing fee.

Characteristics of the Deal

- capital share and ownership share.
- allocation of write-offs.
- financing details, if any.
- obligations of the partners.
- who are the partners.
- any perceived project risks.

11.8 Actual Economic Calculations

The actual calculation of rate of return and payout is calculated as follows:

a) Find gross income per year.

b) Subtract the annual operating cost to find (consider whether any royalty calculations have to be included).

c) Net income per year before taxes.

d) Subtract the depreciation or capital cost allowance allowed for income taxes.

e) The net income before taxes but after depreciation.

f) Subtract the allowable income tax and find.

g) The net income after taxes and depreciation.

h) Add back the depreciation or capital cost allowance to find the cash flow. When the cash flow equals the investment the facility has been paid out.

i) To find the discounted cash flow rate of return, discount the cash flow for each year by an assumed value so that the sum of the discounted cash flow is equal to the investment made.

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12.0 Producing of Gas Well

- Starting a Wellhead: Well Deliverability.
- Design criteria for wellhead equipment.
- Testing of wells.
- Sampling of wells and recombination calculations.

12.1 Introduction

It is necessary to determine how much a gas well will deliver to a gathering system and then to a plant, both initially and on long term, so as to make sure that the contracted quantity of gas is supplied to a buyer. Also it is, of course, necessary to know how much a well will produce so as to size the well site treating equipment. This section discusses the capabilities of wells and how these capabilities are determined. Also, other items such as sampling etc. are discussed.

12.2 Gas Well Equipment

There are many ways of completing a gas well. So that we have a general idea of what happens it is desirable to review what is down the hole in a well. Figure 12.1 shows how a typical well is equipped in the bottom of the hole. The gas enters at the bottom of the casing through perforations and rises through the tubing to the surface. In wells producing sour gas the main part of the casing is sealed off from the gas by means of a packer and the annulus is usually filled with inhibited oil.

The equipment at the wellhead for a sour gas well is shown in Figure 12.2. It primarily consists of two master valves, a flow tee or cross and a wing valve. The Figure shows nearly all the items at the wellhead. Table 12.1 lists all the items. Well heads for low pressure sweet gas wells are equipped with one master valve and screwed valves and fittings are used above the tubing head.

12.3 Well Testing

Nearly all well tests are based on the well known back pressure equation.

Note: Revised to metric by K. Wichert, Petro Canada Ltd.

Equation 12-1

$$Q = C(P_f^2 - P_s^2)^n$$

where

Q = Flow rate in MMSCFD (m^3/d).

P_f = Shut in formation pressure, psia (kPa(abs)).

P_s = Flowing sand face pressure, psia (kPa(abs)).

C = A coefficient that is determined mainly from reservoir characteristics.

n = An exponent that depends primarily on the type of flow in the reservoir at the particular well.

The equation can be plotted on log-log paper, $P_f^2 - P_s^2$ as the ordinate and Q the abscissa. The equation is a straight line with the slope of 1/n. Figure 12.3 shows a typical example. Note that both P_f and P_s are bottom-hole pressures and have to be determined as such, either by direct measurement or by calculation from well head measurements. (Methods are given in the Alberta Energy Resources Conservation Board manual "Gas Well Testing - Theory and Practice", 1979).

Note: The AOF (Absolute Open Flow) is the estimated flow when the pressure at the bottom of the well next to the tubing would be zero and is, of course, a theoretical flow only.

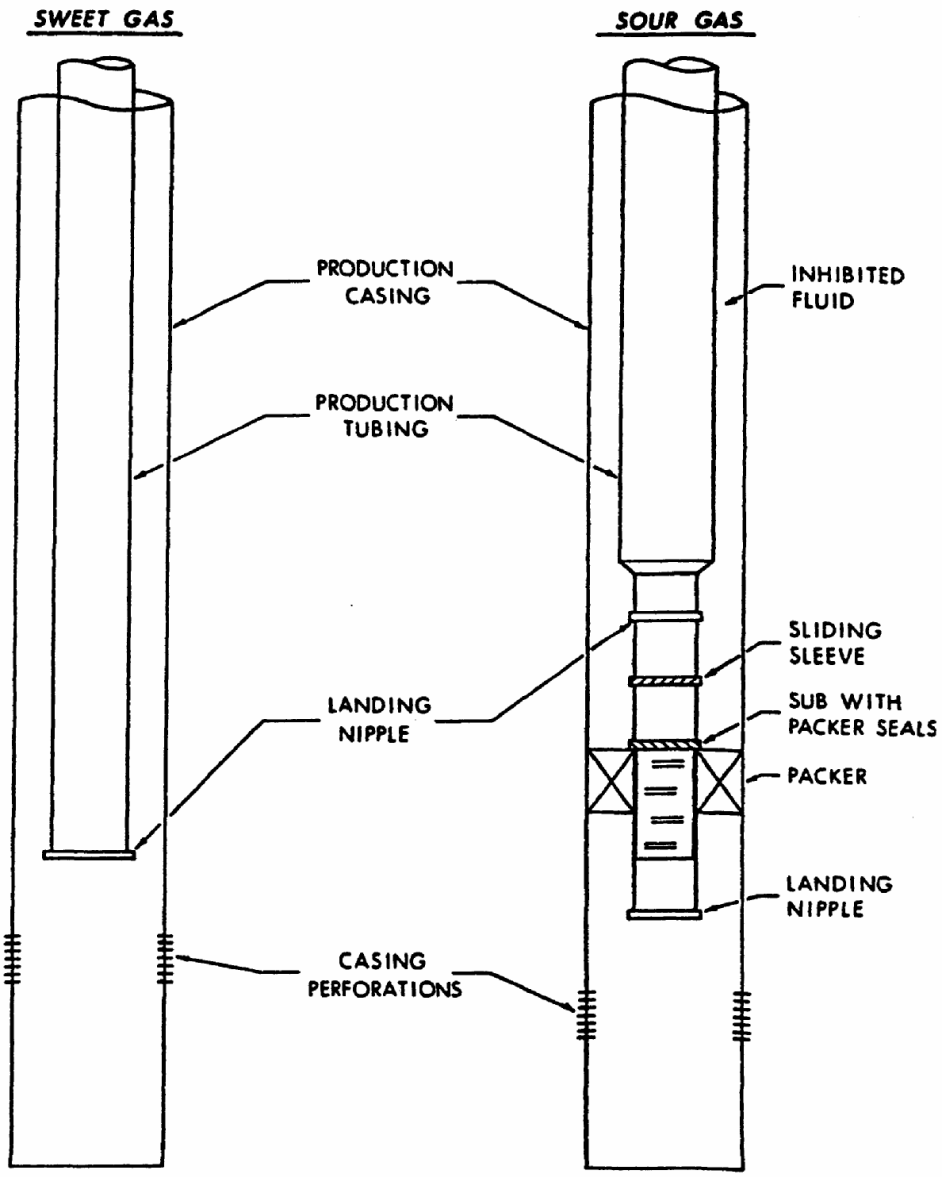


Figure 12.1: Bottom Hole Equipment

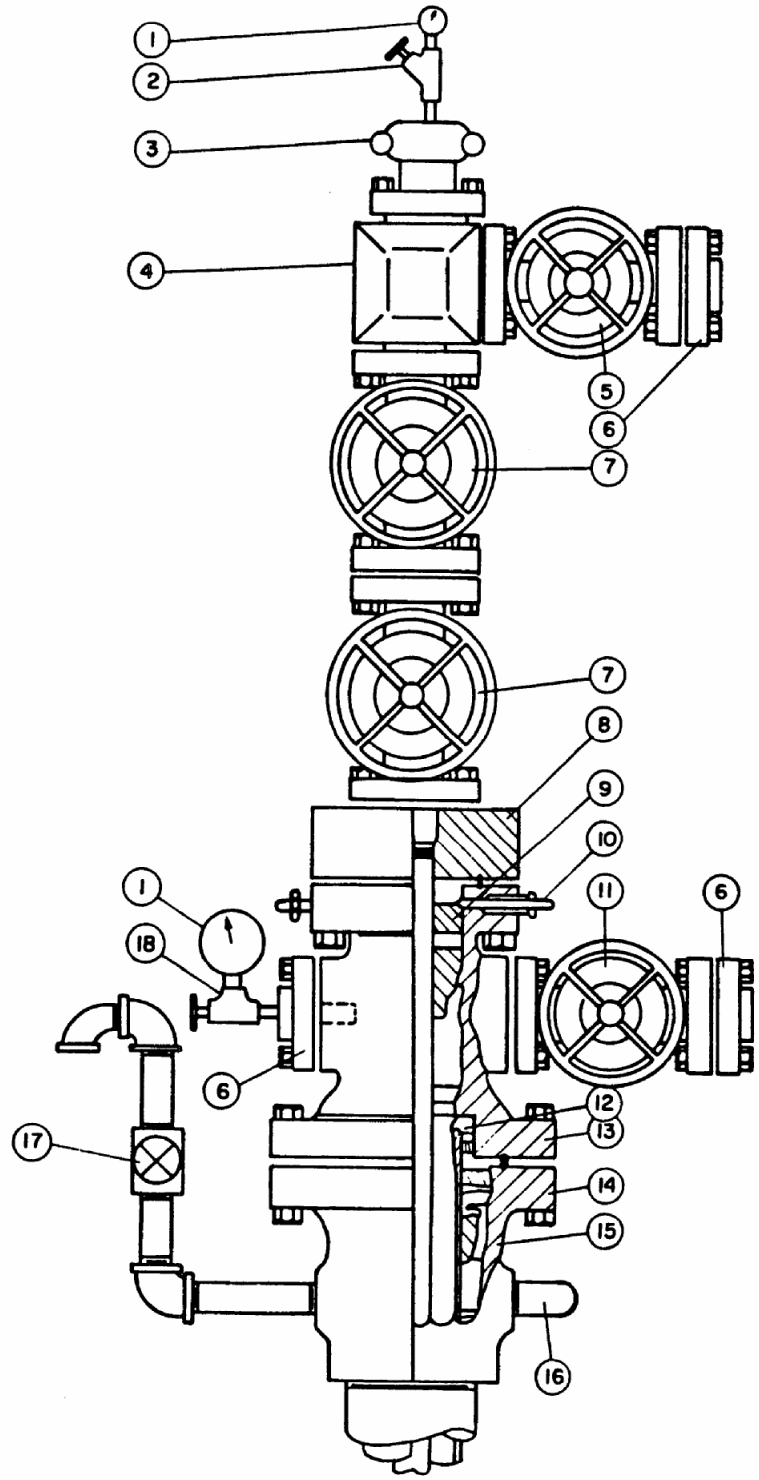


Figure 12.2: Typical Wellhead

Table 12.1: Wellhead Equipment List

Label No. from Figure 12.2	Equipment Description
1	Pressure gauge: 3,000 psi.
2	Needle valve: 1/2 inch - 10,000 pal.
3	Christmas tree cap: 2-1/2 inch, 3,000 psi, tapped 2-7/8 inch EUE.
4	Studded flow tee: 2-1/2 x 2-1/2 x 2 inch, 3,000 psi.
5	Wing valve: 2 inch, 3,000 psi.
6	Companion flange: 2 inch, 3,000 psi.
7	Master valves: 2-1/2 inch, 3,000 psi.
8	Tubing hanger and adapter flange: 6 x 2-1/2 inch, tapped 2-7/8 EUE, 3,000 psi.
9	Wrap around tubing hanger: 7 x 2-7/8 inch.
10	Tubing banger hold down studs.
11	Casing valve: 2 inch, 3,000 psi.
12	Secondary casing pack-off.
13	Tubing head: 10 x 6 inch, 7 inch bowl, with two 2 inch studded side outlets, 3,000 psi.
14	Casing head: 10 x 10-3/4 inch with two 2 inch threaded side outlets.
15	Casing hanger and primary pack-off: 10 x 5-1/2 inch.
16	Bull plug: 2 Inch.
17	Surface casing valve.
18	Needle valve.

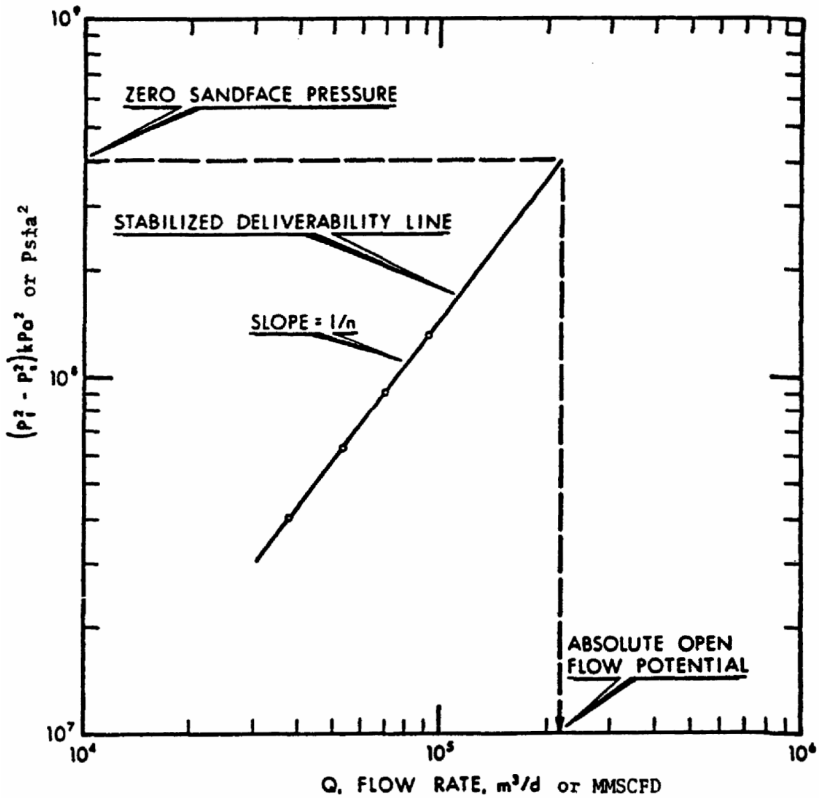


Figure 12.3: Back Pressure Equation Plot

12.4 Types of Tests

There are several methods of obtaining the data for the curve shown in Figure 12.3. The above mentioned manual gives a thorough review of these methods, which briefly are as follows:

12.4.1 Conventional Back Pressure Test

This test is good for wells in formations of high permeability (generally shallow wells). In this test stabilized conditions are obtained. That is, the flow is recorded when the flow pressure is essentially constant. There are various methods for determining when stabilization occurs. One easy method to visualize is when the sand face pressure is reducing at a very slow rate. Sometimes it is considered that at a pressure drop less than 0.1% per 1/2 hour, stabilization is essentially reached. Actually the stabilization time can be very long and other methods have been developed to take care of this situation. These will be discussed shortly.

The method of doing a conventional back pressure test is:

Determine shut-in pressure of formation.

Take successively, in increasing rates, four measures of flow and flowing sand face pressure, making sure that the well is stabilized between each reading. The rates should be adjusted so that a good range of flows is obtained. It is suggested that P_s start at 95% of P_f and go down to about 75% of P_f.

The values of $P_f^2 - P_s^2$ are plotted versus Q on a log-log graph and the four readings joined in a straight line (should be a straight line if steady state conditions exist). The straight line is extrapolated to P_s = 0 to find the AOF of the well.

12.4.2 Isochronal Test

Another type of test is the isochronal test in which flow rates are determined for fixed and equal flow periods, beginning with each flow period at full shut-in pressure. These results are plotted and a value of *n* determined. A fully stabilized test is then run and a line with slope reciprocal of *n* is drawn through it and extrapolated as required.

12.4.3 Modified Isochronal Test

The third test is a modified isochronal in which the shut-in periods are of the same length of time as the flow periods, and P the shut-in pressure, is taken as the pressure at the beginning of the flow test (end of shut-in period) and not the original P_f after complete restabilization.

This data is plotted and "n" determined as usual. Then a full stabilized test is run and the position of the curve determined as was done in the isochronal test.

12.4.4 Modification of Previously Discussed Tests

After wells have been operating some time a single point test is often run. This assumes that *n* is constant from prior tests and only a stabilized test is run.

12.5 Other Data Required During a Test

As the well will be producing into a pipeline it is important to actually obtain a deliverability curve during a test as well as the conventional back pressure curve, although it probably can be calculated from the back pressure data as long as the tubing size and well depth are known.

The deliverability curve uses the equation:

Equation 12-2

$$Q = C'(Psiw^2 - Pw^2)^{n'}$$

where

Q = flow in MMSCFD (m^3/d).

Psiw = shut-in pressure at top of tubing in psia (kPa(abs)).

Pw = flowing pressure at top of tubing in psia (kPa(abs)).

C' = a constant which is determined mainly from the reservoir characteristics of a particular well.

n' = an exponent that depends primarily on the type of flow in the reservoir at the particular well.

Figure 12.4 shows the deliverability curve for a well. As is shown the data at the wellhead is used for this curve rather than the bottom hole conditions.

Besides determining the composition of the gas, which will be discussed shortly, the wellhead temperature, both in shut in and at the various flow conditions should be determined. This is particularly important so as to determine the possibility of hydrate formation at the choke on a pressure reduction.

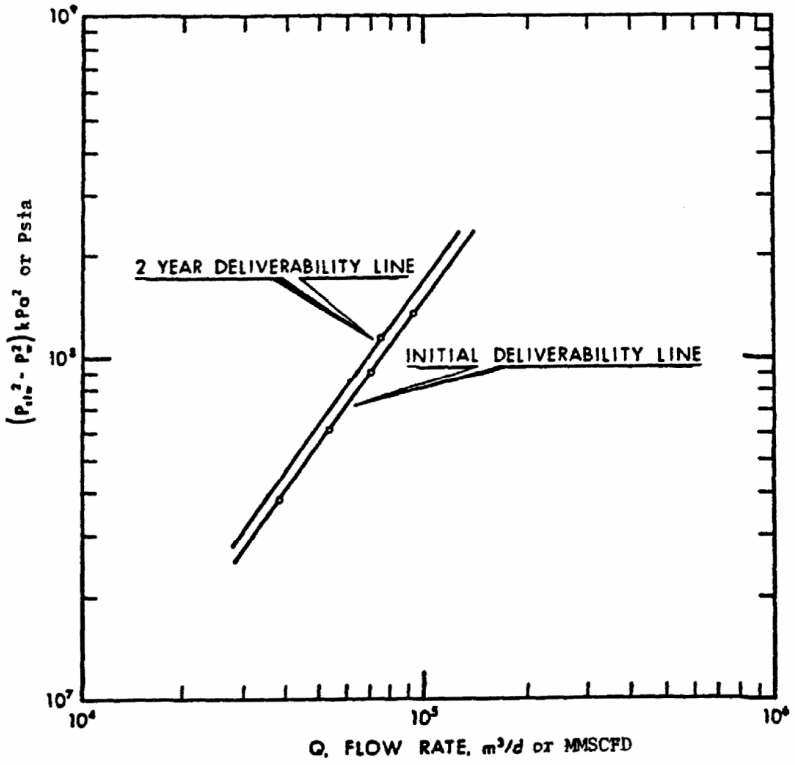


Figure 12.4: Deliverability Curves

12.6 Sampling of Gas During a Test

As the composition of the gas is of prime importance in determining the processing it will get, proper sampling is of utmost importance. Below is given a recommended well test procedure to give good data for determining the composition.

12.6.1 Recommended Gas Well Testing

Tests should be conducted with a standard dual separator hook-up with the high pressure separator operated as close to gathering system pressure as is possible (often about 8 MPa). In addition to the well head pressures and temperatures, the following measurements should be obtained:

- High pressure separator gas rates.
- Low pressure separator gas rates.
- Rate of stock tank condensate production.
- Rate of stock tank vapour production, if possible.
- Rate of water production.

Samples should be obtained at a high production rate after the well has stabilized. Duplicate samples of the following should be obtained:

- High pressure separator gas.
- Low pressure separator gas.
- Low pressure separator liquid.
- Stock tank condensate.
- Stock tank vapours.
- Water.

If it is possible to measure and analyze the stock tank vapour, it will be unnecessary to analyze the low pressure separator liquid.

Each set of hydrocarbon samples should be collected in stainless steel bombs equipped with stainless steel valves. Water may be collected in glass jars. Analytical tests shall include the following:

a) Field Tutwiler tests on the high pressure and low pressure gates (minimum of 10 at ½ hour intervals) and for low H_2S concentration, Draeger or Kitagawa test.

b) On hydrocarbon samples:
   - Standard analysis for N_2, $C0_2$, H_2S and hydrocarbons through to C_7^+.
   - Relative density.
   - Molecular mass and relative density of C_7^+ fractions.
   - Total sulphur content.
   - Quantitative analysis of:
     - Carbonyl sulphide.
     - Mercaptans.
     - Carbon disulphide in the gas samples only.

c) Stock tank condensate
   - Perform distillation per ASTM D216.
   - Salt content.

d) Water
   - Standard water analysis.

12.6.2 Determining the Composition

After the analysis is run as outlined in the previous paragraphs it is necessary to find the actual composition of the well effluent. This is done by recombining the various streams that have been analyzed. Figure 12.5 below shows the flow through the test equipment.

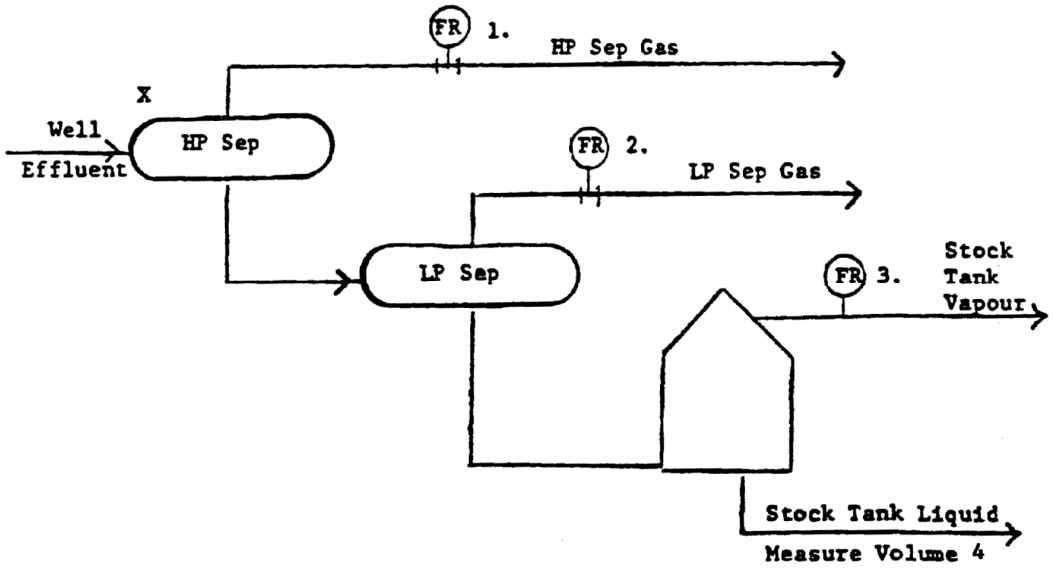


Figure 12.5: Flow through Gas Composition Test Equipment

To find the composition at *X* it is necessary to recombine the fluids measured and analyzed (Points 1, 2, 3 and 4). The form given in Table 12.2 shows how this is done.

Table 12.2: Recombination Calculations

	Stock Tank Liquid			cu .ft. /D	Stock Tank Vapour		Low Pressure Gas		High Pressure Gas		Welihead Fluid	
Comp	LV%	USG PD	cu.ft. /USG		Mole %	cu.ft./D	Mole%	cu.ft./D	Mole%	cu.ft./D	Mole %	cu.ft./D
N_2		*[measure this column]*	91.44			*[measure this column]*		*[measure this column]*		*[measure this column]*		
CO_2			58.48									
H_2S			72.93									
C_1			59.19									
C_2			39.37									
C_3			35.89									
iC_4			29.79									
nC_4			30.87									
iC_5			27.48									
nC_5			27.77									
C_6			21.77									
C_7			17.86									
Total	**100.00**				**100.00**		**100.00**		**100.00**		**100.00**	

The procedure is to analyze the high pressure gas, the low pressure gas, the stock tank vapour and the stock tank liquid f or each individual compound and to measure the total high pressure gas, low pressure gas, stock tank vapour and stock tank liquid, then find the total cubic feet per day at standard conditions of each component, sum them and convert to mole %.

Table 12.3: Recombination Calculations (Metric)

	Stock Tank Liquid			m^3 vap/ d	Stock Tank Vapour		Low Pressure Gas		High Pressure Gas		Welihead Fluid	
Comp	LV%	m^3/ d	m^3 vap/m^3		Mole %	m^3/ d	Mole%	m^3/ d	Mole%	m^3/ d	Mole %	m^3/ d
N_2		*[measure this column]*	682.5			*[measure this column]*		*[measure this column]*		*[measure this column]*		
CO_2			441.6									
H_2S			553.2									
C_1			442.2									
C_2			281.3									
C_3			272.3									
iC_4			229.1									
nC_4			237.6									
iC_5			204.6									
nC_5			206.8									
C_6			182.1									
C_7			162.4									
Total	**100.00**				**100.00**		**100.00**		**100.00**		**100.00**	

The procedure is to analyze the high pressure gas, the low pressure gas, the stock tank vapour and the stock tank liquid for each individual compound and to measure the total high pressure gas, low pressure gas, stock tank vapour and stock tank liquid, then find the total m^3 per day at standard conditions of each component, sum them and convert to mole %.

12.7 The Use of the Data Obtained

The data obtained in gas well sampling and testing is used in many ways. First of all it is used in sizing of well site surface equipment (A. Younger reviews in some detail the data required for proper sizing of gas well surface equipment).

Probably the most important item is what will the well produce. The deliverability curve as shown in Figure 12.4 is an indication of well production capabilities, particularly if the actual pipeline operating pressure is taken as “Ps”. Only the initial deliverability can be obtained by test. Deliverabilities in future years have to be calculated. To do this an estimate of the well completion and reservoir conditions has to be made.

For example, if the pressure is maintained in a cycled reservoir by gas reinjection, then the initial deliverability curve probably applies over the length of the injection period. If the field is produced via the blowdown method, then deliverabilities in future years will probably be considerably lower.

Figure 12.4 shows the calculated deliverability after 2 years. This, of course, assumes that the well has been properly cleaned up on the initial test. Sometimes this is not the case and during the first year the well's performance improves over the course of the year. This can only be determined after a review of the completion program and results.

There are also other methods of determining the design flow rate. These are:

a) 1/4 of the AOF

b) 1/3 of the AOF

c) The Qmax (Maximum Allowable Flow)

Note: a) and b) are arbitrary and have their history in governmental regulations.

Qmax is probably a good value if the reservoir is underlain with water and thus by over-producing this water may be coned up in the well and damage it. However, if the reservoir does not have water underneath, much greater flows can be economically produced. Also, it should be noted that Qmax will decrease with time. The Qmax is calculated according to the Energy Resources Conservation Board of Alberta by the following equation:

Equation 12-3

$$Q\max = qt\left[\frac{P_{FO}^2 - f^2 P_{FO}^2}{P_F^2 - P_S^2}\right]^n \left[\frac{P_F}{P_{FO}}\right]$$

where:

Qmax = suggested allowable flow rate in MMSCFD or ($10^3 m^3/d$).

qt = test flow rate in MMSCFD or ($10^3 m^3/d$).

P_S = the measured flowing sand face pressure during the test.

P_F = the shut formation pressure during the test.

P_{FO} = the original formation pressure.

n = inverse slope back pressure curve.

f = a factor to describe the reservoir. It normally varies from 0.75 to 0.90 depending on the reservoir permeability, porosity, etc.

Other provincial regulatory bodies have similar formulas.

In designing heaters, dehydrators, etc., we need to specify flow rates and estimated wellhead temperature. The wellhead temperature should be measured during the flow test. It should be noted that it will vary significantly between low and high flow rates and it should not be assumed to be constant except at high flow rates. At high rates temperature is higher as heat is not lost as quickly to ground when passing up tubing. Here is a typical flow versus temperature curve for some operating wells:

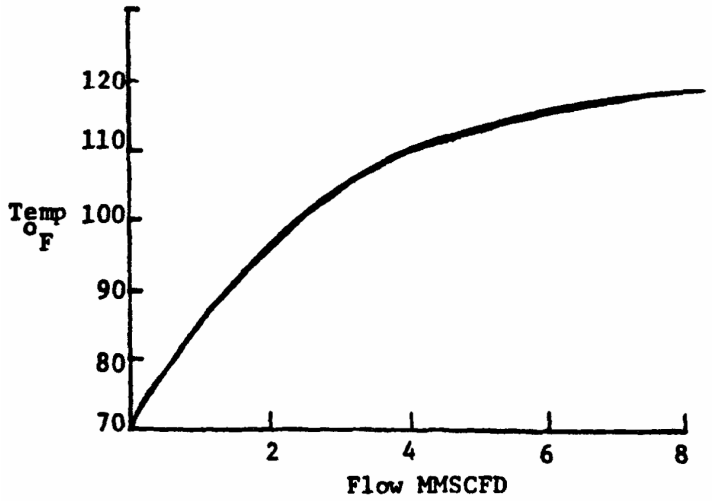


Figure 12.6: Typical Wellhead Temperature Curve

12.8 Problems

Example Problem 12.1

It has been decided that it would be desirable to connect a well that has been tested so that its deliverability against 1400 psia is 10 MMSCFD to an existing gas gathering system operating at 900 psia. What flow should the wellhead dehydrator on this well be sized for?

Additional data

- Assume that the reservoir is not underlain with water.
- The well when it is shut has a pressure of 2200 psia at the surface.
- The AOF based on surface conditions (ie. pressure at surface is atmospheric) is 60 MMSCFD.

References

1. Theory and Practice of Testing Gas Wells, Energy Resources Conservation Board, Calgary, Alberta1 1979.
2. A. H. Younger, J.Can.Pet.Tech., Vol. 7, No. 4, p. 195.

13.0 Field Processing of Gas

- A review of processes used at the wellsite to handle raw gas.

13.1 Introduction

Gas as it leaves the reservoir, is normally saturated with water and, of course, it contains all the impurities normally associated with the gas.

Depending on the composition and the quantity of the gas, it may be treated directly at the well to make it pipeline gas. However, if it is sour and/or there is much liquid in the gas, it is normally only treated so that it can be sent to a central processing facility.

The major problem to handle when transporting the gas is the water content of it and thus the prevention of hydrates. For hydrates to form the gas temperature must be below the temperature of the hydrate formation and at a pressure greater than the hydrate pressure and there must be free water present.

Three methods are normally used to handle gases at the well head:

1. Heating the gas to keep it at all times above the hydrate temperature.
2. Removing the water from the gas so that no free water will be formed by cooling in the transmission line and thus no hydrates formed.
3. Combining the free water formed with some chemical so that a hydrate cannot form.

Figure 13.2 and Table 13.2 summarize the alternates.

13.2 Heating

There are two methods used for heating. The most common is the use of well heaters and line heaters spaced as required. The other method used is hot water tracing.

13.2.1 Heaters

Figure 13.1 shows a typical wellhead heater. The gas is produced from the well through a set of coils to preheat the gas to take care of the temperature drop that occurs on choking the gas. It then passes through the choke and the rest of the heating coils until it is heated to the desired outlet temperature. This temperature is controlled by regulating the fuel to indirect burners in the heater. Some additional points to consider are:

1. Heat Transfer Fluid - This is normally a mixture of ethylene glycol and water. The mixture should be such that problems will not occur if the unit is shut down for an extended period of time. This does not mean that the solution does not freeze but only that a dangerous solid ice phase does not exist. For example, a solution that is good for -40ºF (-40ºC) operation is 25% by weight of Ethylene glycol which freezes at about -10ºF (-25ºC) but only forms a mush type of mixture.

   Sometimes straight water is used as it gives a better heat transfer coefficient but particular care has to be taken when the unit is shut down. Also, sometimes a very low pressure steam is generated as this gives even better heat transfer coefficients.

2. Gas Outlet Temperature - The temperature of the gas out of the heater generally should be set by the maximum allowable temperature that the pipeline coating can stand. Some recommended outlet temperatures are given in Table 13.1.

   It should be noted that particular care should be taken to design the pipeline for proper temperature expansion allowance when using heaters.

Table 13.1: Allowable Heater Outlet Temperatures

Pipeline Coating	Maximum Heater Gas Outlet Temperature ºF	
Yellow Jacket Polyethylene	165º	(75ºC)
Blue Guard Polyethylene	165º	(75ºC)
Poly Vinyl Chloride	165º	(75ºC)
Coal Tar Products	200º	(95ºC)

3. Specifications - The items that should be specified when purchasing heaters are as follows:

   a) The gas composition.

   b) The operating pressure for the heater.

   c) The wellhead pressure. This is needed for sizing the choke and the sizing of the heating coil upstream of the choke.

   d) The wellhead temperature. This information is needed as in c). Also it is needed for sizing the heater.

   e) The reservoir pressure and temperature. This gives the amount of water that may condense out after the choke. It is normally not too significant.

   f) The gas outlet temperature. This, along with c), d) and i), determines the size, ie. BTU rating of the heater.

   g) The heat transfer fluid to be used. This was discussed previously.

   h) The pressure drop allowed. This determines tubing size in the heater. Normally should be between 5 to 10 psi (35 to 70 kPa).

   i) The design flow. The expected flow through the heater is to be used (See previous chapter). For sizing line heaters please see later in the chapter on what to use.

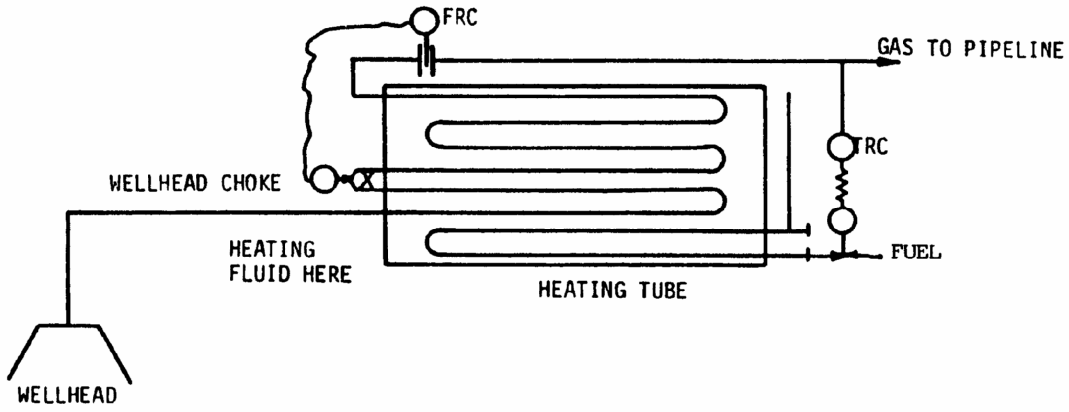


Figure 13.1: Typical Heater Installation

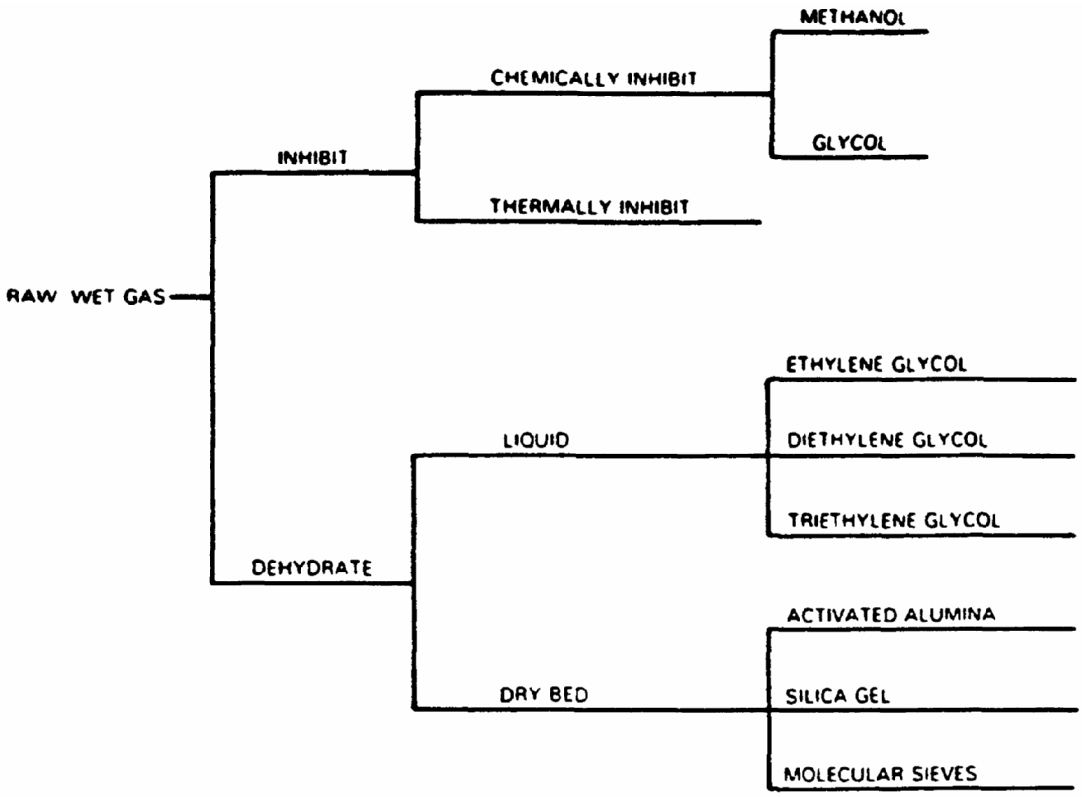


Figure 13.2: Options for Hydrate Control

Table 13.2: Hydrate Control Methods

Method	Variation	Positive Features	Negative Features	Major Applications
Chemical inhibition	Methanol	Inexpensive equipment Easy to handle	High methanol losses Introduction of a potential contaminant	Seasonal control in gathering systems Temporary control in cryogenic plants
	Glycol	Inexpensive equipment High recovery and reuse rate	Difficult to handle in very low temperature conditions May reduce efficiency of downstream heat exchangers	Gathering systems where regeneration equipment is unavailable or undesirable Low temperature plants for localized control
Thermal inhibition (heat)	Steam or electric tracing of pipes Gas heaters in field	Ease of control Good for sour gas wells	None	None
Liquid (Glycol)	None	Relatively inexpensive equipment Easy to operate and maintain Energy consumption fairly low in proportion to the water load	Limited dew-point depression Subject to foaming in some gas systems	Gathering systems, wellheads Feed to transmission lines Feed to moderately low temperature processes
Dehydration	Dry bed	Equipment reliable for unattended operation Very low dew points are possible Can be designed to dry and	Equipment cost is higher than other kinds of dehydrators Energy requirements are relatively high and not	Feed to cryogenic plants Liquid hydrocarbon drying Combination systems Hydrocarbon dew-point

		sweeten simultaneously, or to control hydrocarbon dew point	completely in proportion to the water load	control

13.2.2 Hot Water Tracing

Another method of hydrate control is that developed by Petrofina Canada Ltd. in the Wildcat Hills Field. This method consists of running hot water lines next to the pipe to keep it warm. A circulating water system with a hot water generator (several are used) is reported to be quite successful, but somewhat costly to install. A good reference for describing this process is by Lukacs and Nader.

Besides Wildcat Hills, the Clarke Lake Field at Ft. Nelson and part of the Pine Creek Field near Edson are operated using this system.

13.2.3 General

Both the heater system and the hot water tracing system do not remove any water in the treatment system. All water that is produced at the wells is carried to the plant and has to be disposed of there. Also, since on choking free water is formed there will nearly always be water flowing along the bottom or sides of pipe (depends on type of two phase flow). This sounds like it could be a problem in sour gas systems. However, as has been mentioned, there are several systems that are operating very successfully without reporting corrosion problems.

13.2.4 Design

The problem of designing systems such as heater or hot water tracing systems is actually calculating the loss of heat from the buried pipe and making sure the fluid flowing through it does not cool to a temperature below the hydrate formation temperature of the fluid. The length of line that can be used for reheating or how much the gas has to be heated can be calculated from the following equation:

Equation 13-1

$$q = UA\frac{(T_1 - t) - (T_2 - t)}{\ln\left(\frac{T_1 - t}{T_2 - t}\right)} = UA(\Delta t_{lm})$$

where:

U = the overall heat transfer coefficient for transfer from buried pipes to the ground BTU's/HR/sq ft/ºF (W^2/m^2/HR/ºC)

A = the area of transfer, ie. pipe perimeter *x* its length ft^2 (in^2)

T_1-t = the initial temperature (T_1) - outside ground temperature (t). The final temperature is some temperature above hydrate point; usually about 5ºF (3ºC) above the hydrate temperature (see Chapter 5 for a discussion on hydrates).

An easier way is to use enthalpies at the pressure and temperatures of the system, ie. $q = W(H_1 - H_2)$ then use equation 1 for finding A. The overall heat transfer coefficient, U's value, varies considerably with the soil conditions. Table 13.3 gives some experimental determined values.

Temperatures expressed in ºF or ºC as appropriate.

q = the rate of heat transfer at steady flow rates BTU's/hr (w) and by the following equation:

Equation 13-2

$$q = W\left(H_1 - H_2\right)$$

where:

W = the rate of flow, lbs/hr or Kg/hr

H_1 = the enthalpy per unit of mass at temperature T_1 and the pressure at 1.

H_2 = is the enthalpy per unit of mass at temperature T_2 and the pressure at 2.

Combining equations 1 and 2 we get

Equation 13-3

$$A = \frac{W\left(H_1 - H_2\right)}{U\left(\Delta t_{lm}\right)}$$

and thus we can solve for A.

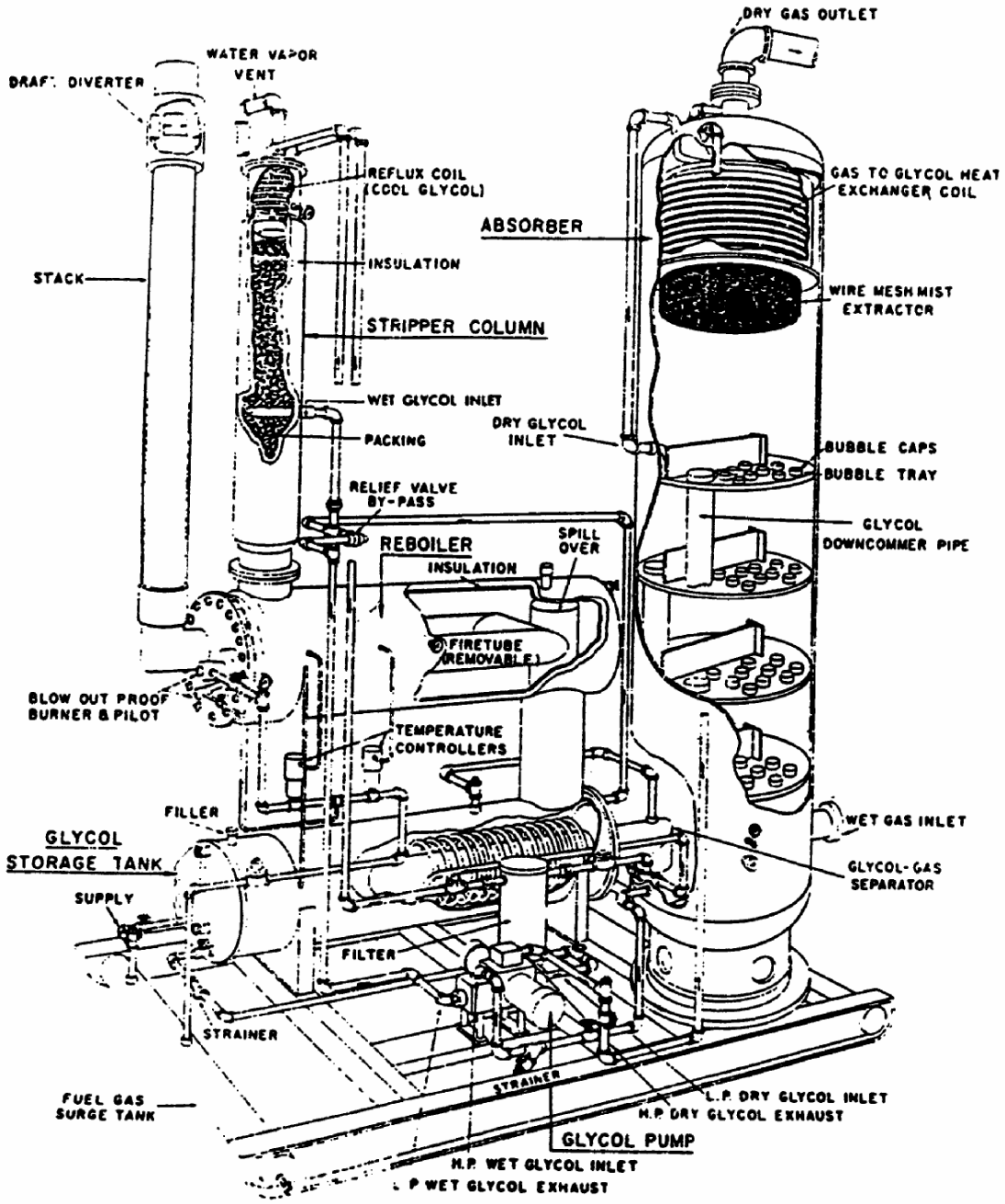


Figure 13.3: Glycol Dehydrator Components

Table 13.3: Typical Ground 'U's'

Soil	BTU/hr/ft^2/^{o}F	W/m^2/^{o}C
Clay and Gravel	0.618	3.51
Shale and Muskeg Mix	0.500	2.84
Muskeg (Dry)	0.400	2.27
Muskeg (Wet)	2.000	11.30
Silty Gravel	0.600	3.41

A good design safe value is 0.65 BTU/hr/ft^2/ºF (3.7 w/m^2/ºC). If insulated pipe is used a design value of 0.15 BTU/hr/ft^2/ºF (0.85 w/m^2/ºC) for overall heat transfer coefficient can be used. Field tests show that the actual values generally run a little lower than this.

13.2.5 Line Heaters

In selecting the need and number of line heaters the expected design flow should not be used, but the normally expected minimum flow should be used or ½ the design flow to determine the location or need of a line heater. For example, say that a wells deliverability is 30 MMSCFD. To calculate where to locate a line heater or if it is necessary on the transmission line, a flow of 15 MMSCFD should be used in equation (3) and solved for area of pipe, and thus the length before a hydrate can be expected to occur.

13.3 Water Removal or Dehydration

13.3.1 Glycol Treaters

This is probably the most common method of water removal. A flow sketch as given in Figure 13.4 and Figure 13.3 shows the facilities used.

The glycol contacts the gas in a small trayed (4-6 trays) or packed tower and absorbs the water vapour. The dried gas goes into the pipeline and the hydrocarbon liquids are separated in the scrubber and recombined with the gas.

The glycol and water solution is reduced in pressure, some flash gas let off and the water stripped out of the glycol in a still. The glycol is then recirculated by a pump using normally high pressure gas for a driving mechanism. Generally a slipstream of glycol is filtered after it has been stripped.

Various glycols can be used; generally diethylene glycol is used but sometimes triethylene glycol is used. The type used is generally set up by the amount of dehydration required. This is set by the inlet gas temperature and the minimum temperature that will be expected in a pipeline. In Canada, the minimum temperature is generally set at 25ºF (-4ºC). Therefore, if the inlet gas temperature is 80ºF (26.5ºC), then a dewpoint reduction of 55ºF (30.5ºC) is required. Table 13.4 gives some pertinent information regarding both diethylene and triethylene glycol in dehydration service.

Table 13.4: Glycol Data

Item	Diethylene Glyco	Triethylene Glycol
Normal strength with straight dehydration %	95	98.5
Regeneration temperature ºF (ºC)	330 (165)	360 (182)
Dewpoint Suppression ºF (ºC)	65 (36)	90 (50)
Dewpoint Suppression with Gas Stripping ºF (ºC)	110 (61)	140 (78)
Degradation or breakdown temperature ºF (ºC)	350 (177)	403 (206)

Note: Charts on Page 20-20 and 20-20 of the 1987 English Unit GPSA Data gives more information on this.

Some design criteria are as follows:

- The normal circulation rate for these units is about 4 gals of glycol circulated per lb. of H_2O to be removed (about 30 L glycol per Kg of H_2O picked up. The glycol contactor

usually has 4-6 trays or the packing equivalent of same. Most of these towers are vertical however; sometimes special horizontal packed vessels are used.

- The regeneration unit should be sized for 2000 BTU/gallon (560 kj/l) of liquid circulated, providing that good heat exchange prior to the actual regenerator is provided, higher if it is not. A good article on dehydration unit design is given by L.S. Ried.
- Some of the operation problems with these units are:
- Rotating equipment operating unattended for several hours sometimes gives problems.
- Glycol absorbs some gas, particularly H_2S. It is believed to be the start of pollution problems in the Pincher Creek area.
- Glycol consumption can be quite high and thus costly; a design figure is 3-4 lbs/MMSCF (50-65 kg/10^6m^3) with sour gas, but has run 10 times this value in some places. For sweet gas it is generally less than 1 lb/MMSCF (15 kg/10^6m^3).
- In condensate gas fields, the hydrocarbon is not treated and sometimes absorbs or occludes water which is then vapourized back into the gas and hydrates do occur down the pipeline,
- Glycols used in field locations can become contaminated with corrosive compounds or heavy oil that cause foaming and other problems.
- Oxygen or air, if allowed to enter the system through the storage tank, sump tanks or badly leaking packing glands will cause oxidation of the glycol. This creates organic acids that are very corrosive. A gas blanket on the storage facilities and good pump maintenance will prevent most of this.
- Every effort should be made to keep free water from entering the dehydrator. his free water most often is a brine. The brine concentrates in the regenerator until such time as salt drops out. It will scale up the reboiler tubes reducing heat transfer and very often resulting in a hot spot and possibly burning out of the fire tube. Salt brine in glycol solutions also changes the specific gravity of the glycol so that erroneous strengths based on gravity are obtained. Also brine, if associated with air at any point, becomes extremely corrosive.

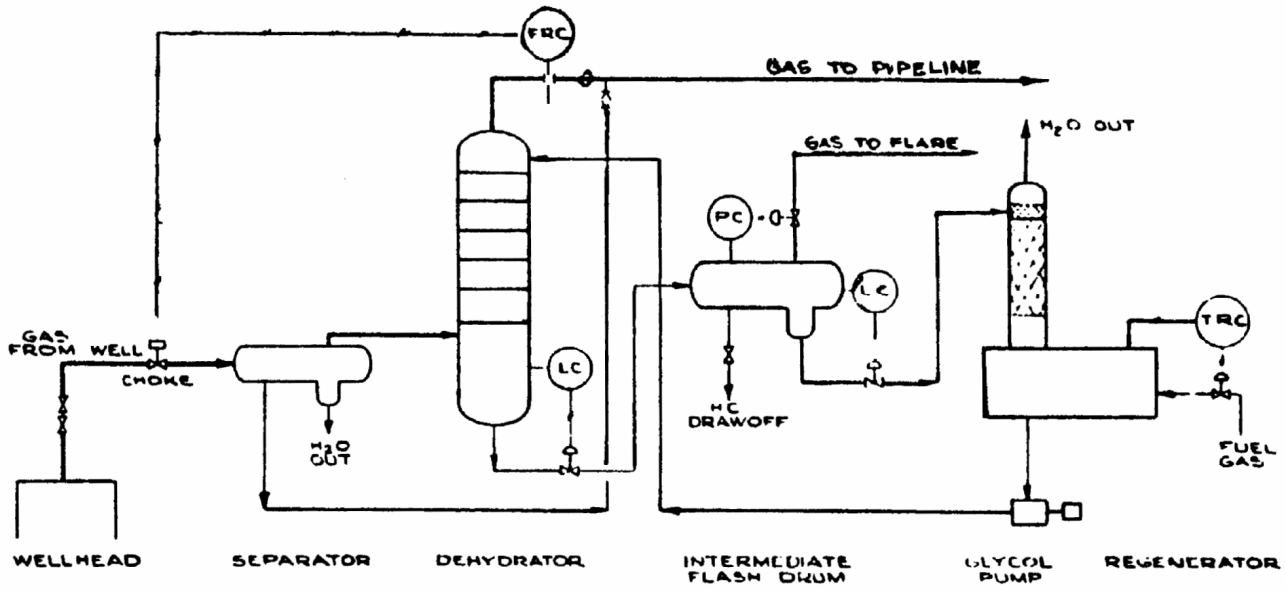


Figure 13.4: Typical Dehydrator

13.3.2 $CaCl_2$ for Water Removal

$CaCl_2$ as a solid spontaneously absorbs water (deliquescence). Also, the concentrated solution of it absorbs water and these solutions normally have a lower vapour pressure of water than that prevailing in the atmosphere at the same time. Figure 13.5 shows a dehydrator in some detail. It consists of passing the gas through a tower filled at the top with solid $CaCl_2$ and then some liquid trays in which the calcium chloride solution flows across and the gas bubbles through so that additional water can be absorbed at the bottom where the gas enters a brine discharge section.

The expected usage of $CaCl_2$ is about 1 lb. $CaCl_2$ per every 3.5 lbs of water (1 kg $CaCl_2$ per 3.5 kg of water). The expected water content of gas leaves this unit in usually less than 2 lb/MMSCFD of 1,000 psig gas (30 kg/$10^6 m^3$ of gas at 7 MPA). Some people claim an economic advantage to the use of these units (see Fowler). However, there are several

problems:

1. Does not work when have emulsified oil in gas.
2. Have to figure out how to dispose of $CaCl_2$ brine.
3. Hard to recharge - need considerable recharging.
4. Solid $CaCl_2$ bridges badly.
5. The saturated solution freezes quite easily and causes many problems.

A particularly good application is small sweet gas wells such as wells producing from the Milk River or Viking gas zones, ie. for wells producing 0.500 MMSCFD or less 15,000 m^3/day or less).

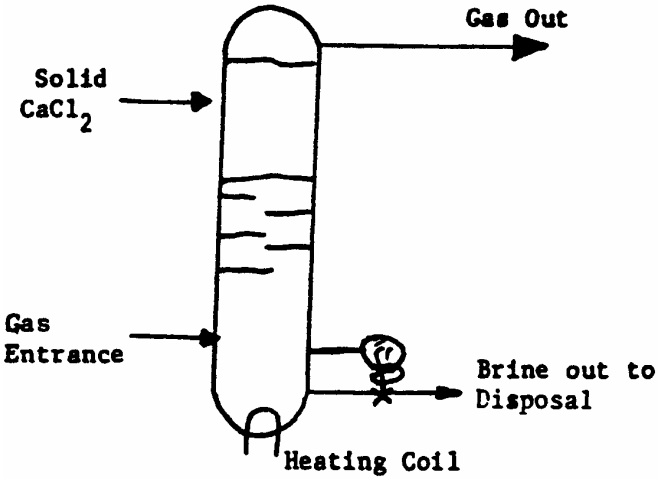


Figure 13.5: $CaCl_2$ Dehydrator

13.3.3 Specifications

The data required for a specific dehydrator is as follows:

- The gas composition.
- The operating pressure.

- The wellhead pressure and temperature. This data is needed for sizing the choke and the separator just ahead of the contactor.
- The reservoir pressure and temperature. This is required to see how much water is in the gas entering the unit.
- The dewpoint of the gas out of the unit - usually about 25ºF (-4ºC).
- The design flow rate.
- The pressure drop allowed through the unit. Enough should be allowed so that the hydrocarbon liquid formed in the separator can be by-passed properly. About 25 psi (170 KPA). is often allowed.

13.4 Combining the Free Water with Chemicals

Another method of hydrate control is the addition of an absorbing chemical, normally methanol, to the gas stream. In this case the methanol is lost and this method is quite expensive on a continuing basis and is used as an emergency protection or to remove hydrates when they form.

Glycol injection, as practiced in the gas plant, has been tried and worked reasonably successful in the field in which there is not a large liquid phase or any sour gas. Withers and Leffingwell describes a successful system.

The amount of glycol or methanol required can be calculated from the well known Hammerschmidt equation:

Equation 13-4

$$d = \frac{KW}{100\,M - MW}$$

where:

d = the dewpoint suppression required.

M = Mole weight of alcohol.

W = Weight percent of alcohol in final solution.

K = a constant, equals 2335 (1297) for methanol, ethanol, ammonia – 4000 (2222) for any glycol.

Note: The example problem 13 in Chapter 5 illustrates the circulation required.

13.5 Special Methods

13.5.1 Short Cycle Units

Sometimes it is desirable to also recover hydrocarbons. In this case a short cycle adsorption unit as described in Chapter 17 is used.

13.5.2 Low Temperature Dewpoints or High Water Content Gases

For very low temperature dewpoints or for very high water content gases, special dessicants such as molecular sieves, alumina, etc. are used.

Figure 13.6 shows a typical solid bed dessicant unit. The beds absorb water for 6-8 hours then are regenerated and the other put on line. Table 13.5 below gives performance data on various dessicants used. For cryogenic plants where water contents of less than 0.1 ppm are required molecular sieves are generally used. Page 20-22 to 20-24 of the 1987 English Units GPSA Data book gives details of the design of molecular sieve units.

Table 13.5: Typical Desiccant Properties

Desiccant	Shape	Bulk Density (lb/ft^2)	Particle Size	Approximate Minimum Moisture Content of Effluent Gas (ppmw)
Alumina Gel	Spherical	52	1/4"	5-10
Activated Alumina	Granular	52	1/4"-8 mesh	0.1
Activated Alumina	Spherical	47-48	1/4"-8 mesh	0.1
Silica Gel	Spherical	50	4-8 mesh	5-10
Silica Gel	Granular	45	3-8 mesh	5-10
Mole Sieve	Spherical	42-45	4-8 mesh	0.1
			or 8-12 mesh	0.1
Mole Sieve	Extruded Cylinder	40-44	1/8"	0.1
			or 1/16"	0.1

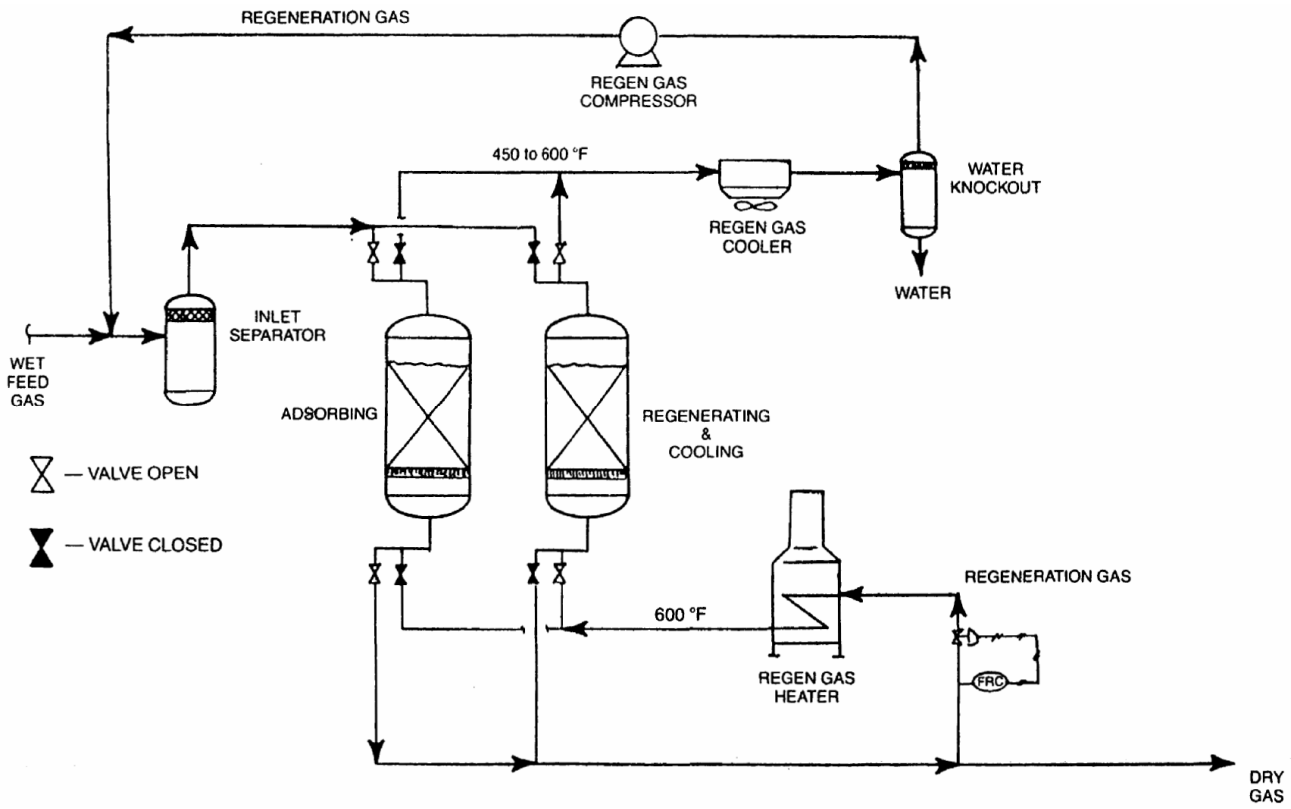


Figure 13.6: Example Solid Desiccant Dehydrator Twin Tower System

13.6 Problems

Example Problem 13.1

It has been decided to use heaters for hydrate control in sweet gas fields. You are to size the heater for a typical well in MMBTU's per hour.

Additional data

Gas gravity	0.7
Welihead temperature	130ºF
Wellhead flowing pressure	2000 psi
Deliverability at 2000 psia	15 MMSCFD
Deliverability at 1500 psia	20 MMSCFD
Wellhead shut-in pressure	3000 psia
Pipeline normal operating pressure	1000 psia
Hydrate temperature of gas	65ºF
Pipeline coating	yellow jacket
Line size to plant from well	6 inch
Soil	Shale and muskeg mix
Length of line to plant	4 miles
Area of outside surface of 6 inch pipe per foot of length	1.734

Also please indicate by calculations if you believe it is necessary to install a line heater on this line if this is the only well on the line to the plant. The area of the field is Alberta, Canada.

Example Problem 13.2

A well that is two miles from a processing plant in Alberta produces gas to a plant without any heating or dehydration. It has been decided to set a minimum flow alarm on the well to warn the plant operators when the line might be close to having hydrates forming. Determine what this flow is in MMSCFD.

Additional data

- The wellhead temperature after choking is 100ºF.
- The line to the plant is 6" diameter (the area per lineal foot of 6" pipe is 1.734 ft^2) - ordinarily coated pipe is used.
- The soil to the plant is silty gravel.
- The gas mole weight is 20.
- The gas specific heat at the operating pressure is 0.64 BTU's/lb.
- The hydrate temperature of the gas is 68ºF.

Example Problem 13.3

A shallow gas gathering system has 25 wells that have a capability of producing 5 MMSCFD. It is decided that the method of controlling hydrates and/or freezing of the gathering lines in the winter is to inject methanol at the wellhead. Calculate how much methanol has to be injected in a 4 month period.

Additional data

- The wellhead temperature is 40ºF.

- Average gathering system pressure is 300 psi.
- Gas water content in reservoir is 28 lbs/MMSCFD.
- Average ground temperature is 25ºF.
- The methanol used is 95% pure with 5% H_2O.
- The methanol is not recovered.

References

1. J. Lukac and W. Nader, Journal of canadian Petroleum Technoloqv, Summer 1964, P. 7.
2. L.S. Ried, Journal of Petroleum TechnoloQv, Aug. 1970, P. 925.
3. O.W. Fowler, Oil & Gas Journal, Aug. 3, 1964.
4. B.M. Withers and J. Leffingwell, Oil and Gas Journal, Feb. 6, 1967, P. 82.
5. D. Ballard, "The Fundamentals of Gas Dehydration," 1979 Gas Conditioning Conference at Norman, Oklahoma.

14.0 Transporting of Fluids to the Plant

- A review of pipeline designing not covered in fluid flow.

14.1 Introduction

All fluids are carried to the plant in steel pipes. The use of other types of material for gas gathering facilities is very limited, plastic has been used in some cases in gas distribution lines at low pressures and coiled aluminum has been used for high pressure lines.

14.2 Determination of Size

The detailed determination of size has been covered in the section on fluid flow (see Chapter 6). We will just review the more pertinent items.

14.2.1 Design Flow

The design flow in each branch of the gathering system is calculated backwards from the processing plant allowing for the shrinkage and fuel use in the plant and at the wellhead. Each line in the gathering system should be investigated to decide what the maximum flow is down the branch, taking into account the maximum anticipated flow from the wells supplying the branch. The problem given below illustrates this calculation.

Example Problem 14.1

List the design flow rates for the gathering system shown in Figure 14.1. The raw gas composition is as follows:

Compound	Mole %
N_2	2.79
H_2S	11.42
CO_2	2.93
C_1	74.43
C_2	3.18
C_3	1.67
iC_4	.35
nC_4	.89
iC_5	.33
nC_5	.35
C_6^+	1.66

Note: This Chapter revised to metric by E. Wichert.

The products from the plant will be residue gas, sulphur, pentanes plus containing up to 8% butane. Fuel used is estimated for both the plant and field to be approximately equivalent to 2% of the raw

gas to the plant, but of course sales gas is used. The maximum production of residue gas will be 25.0 MMSCFD $\left(700\,\frac{10^3 m^3}{d}\right)$. The design deliverabilities for each well connected are as follows:

Well No.	Design Flow at 1200 psi	(8 MPA)
	MMSCFD	(10^3m^3/d
11-14	10.0	(280)
10-22	4.5	(125)
10-23	5.5	(155)
11-28	6.0	(170)
10-27	4.0	(110)
6-25	2.0	(55)
10-19	1.5	(40)
10-10	6.5	(185)

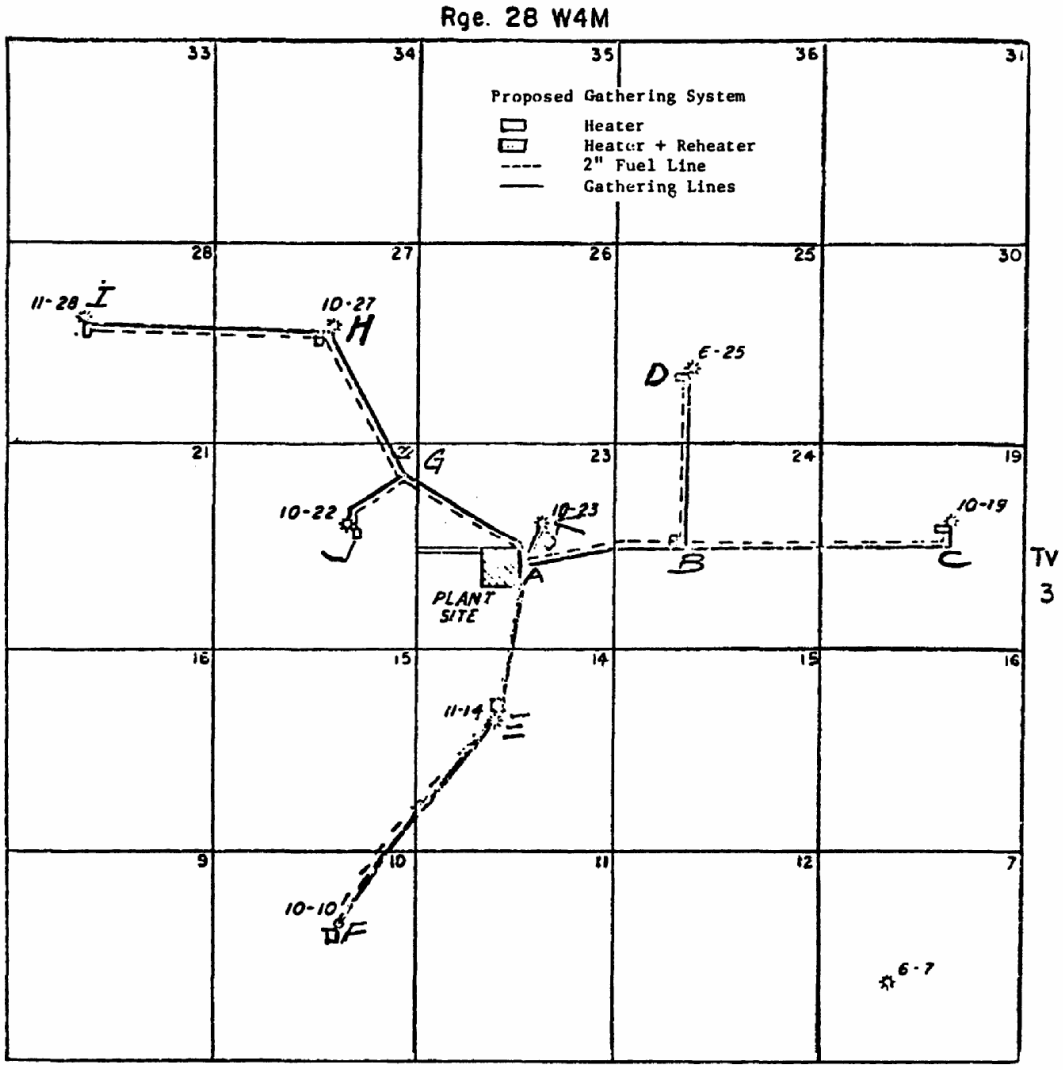


Figure 14.1: Typical Gathering System

Solution

Calculate the raw gas required at the plant entrance.

Shrinkage for acid gas	14.35%
Shrinkage for pentanes	2.34
Shrinkage for butane	0.2
Sub-total	16.89
Fuel gas	2.00
Total Shrinkage	18.89

Raw gas to be delivered to the plant:

$$\frac{25.0}{0.8111} = 30.8\ MMSCFD \quad or \quad \left(\frac{700}{0.8111} = 863\ 10^3 m^3/d\right)$$

When determining the deliverability required by the wells, add at least

10%. Therefore the well deliverability should be approximately 34 MMSCFD minimum. It is actually 40 MMSCFD (1120 $10^3m^3/d$) in the case.

Line Description	Max. Flow Expected	Design Flow
A-B	2.0 + 1.5	3.5**
B-D	2.0	2.0
B-C	1.5	1.5
A-G	6.0 + 4.0 + 4.5	14.5**
G-J	4.5	4.5
G-H	6.0 + 4.0	10.0
H-I	6.0	6.0
A-K	5.5	5.5**
A-E	6.5 + 10.0	16.5**
E-F	6.5	6.5

** Sum of lines to plant 14.5 + 16.5 + 5.5 + 3.5 = 40.0 MMSCFD compared to 30.8 MMSCFD required. Probably should design for the 40 MMSCFD or at least 1.1 (30.8) = 34 rather than the 30.8 required.

Metric Answer

Line Description	Max. Flow Expected $10^3m^3/d$	Design Flow $10^3m^3/d$
A-B	55 + 40	95**
B-D	55	55
B-C	40	40
A-G	170 + 110 + 125	405**
G-J	125	125
G-H	170 + 110	280
H-I	170	170
A-K	155	155**
A-E	185 + 280	465**
E-F	185	185

** Sum of lines to plant 95 + 405 + 115 + 465 = 1120 $10^3m^3/d$ compared to 863 $10^3m^3/d$ required. Probably should design for the 1120 $10^3m^3/d$ rather than the 863 $10^3m^3/d$ required.

14.2.2 Allowable Pressure Drop

The allowable pressure drop for the gathering system is also determined by working backwards from the process plant. The normal maximum plant delivery pressure is set at 900 psia. Then after assuming a 50 to 75 psi pressure drop through the processing plant, the terminal pressure of the gathering system may be determined. The starting pressure is generally that of a 600 ASA flange 1440 psig (10024 KPa) or if pipeline thickness dictates slightly lower pressures.

Note: The full pressure drop is not often utilized for pressure drop from the wellsite control valve to the plant entrance. Pressure drops of 20 psi per mile (135 KPa per mile) are often used and with wells that are close to the plant, possibly up to 50 psi per mile (348 KPa per mile).

14.2.3 Calculation of Pressure Drop

The pressure drop in the gas gathering is calculated by the general gas pressure drop equation (metric given on next page):

Equation 14-1

$$Q = \frac{5634 d^{2.5}\left(T_b + 460\right)}{P_b\left(STlZ\right)^{1/2}} \frac{1}{\sqrt{f}} \left(P_1^2 - P_2^2\right)^{1/2}$$

where:

Q = Flow in SCFD at the base conditions of T_b and P_b.

d = Pipe diameter in inches.

ℓ = length in feet.

P_1 = Pressure upstream in psia.

P_2 = Pressure downstream in psia.

Z = Compressibility factor.

S = Specific gravity of the gas as compared to air.

f = The friction factor as obtained from Figure 14.2.

T = The operating temperature in ºF or in metric units.

Metric

$$Q = 0.1034 \frac{d^{2.5}}{\left(GTLZ\right)^{0.5}} \frac{1}{\sqrt{f}} \left(P_1^2 - P_2^2\right)^{0.5}$$

where:

Q = Flow in m^3/d.

d = Pipe diameter in mm.

L = length in meters.

P_1 = Pressure upstream in KPa (abs).

P_2 = Pressure downstream in KPa (abs).

Z = Compressibility factor, at $(P_1 + P_2)/2$ and T.

G = Relative density of the gas (air = 1).

f = The friction factor as obtained from Figure 14.2.

T = The operating temperature in ºK.

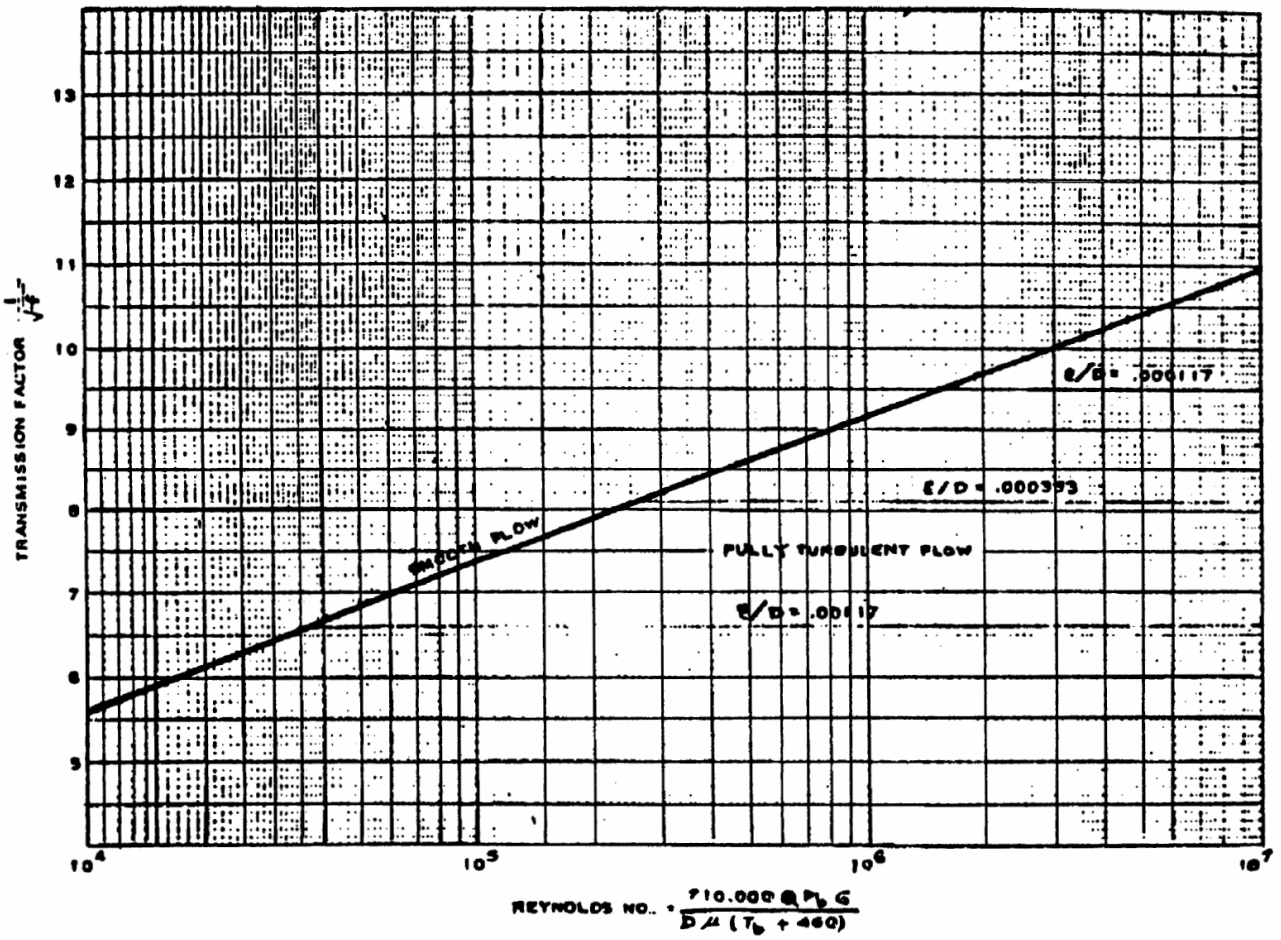


Figure 14.2: Recommended Friction Factors for Gas Flow

The pressure drop is also calculated sometimes by use of either the Weymouth or the Panhandle formulas. In these equations, which are given below, the friction factor has been worked into the constant of the equations.

Weymouth Formula

Equation 14-2

$$Q_D = 433.5 \frac{T_b}{P_b} \left[\left(\frac{P_1^2 - P_2^2}{STlZ} \right) d^{16/3} \right]^{1/2}$$

where:

d = pipe diameter in inches.

Q_D = flow in SCFD at the base conditions of T_b and P_b.

ℓ = length in miles.

T = operating temperature in ºR.

T_b = temperature base ºR.

P_b = pressure base in psia.

G = specific gravity.

Z = average compressibility.

Panhandle Formula

Equation 14-3

$$Q_D = 737\left(\frac{T_b}{P_b}\right)^{1.02} E\left[\frac{P_1^2 - P_2^2}{G^{0.961} LTZ}\right]^{0.51} d^{2.53}$$

where:

d = pipe diameter in inches.

Q_D = flow in SCFD at the base conditions of T_b and P_b.

ℓ = length in miles.

T = operating temperature in ºR.

T_b = temperature base ºR.

P_b = pressure base in psia.

G = specific gravity.

Z = average compressibility.

E = the pipeline efficiency which is generally assumed to be about 88%.

The metric versions of these equations are:

Weymouth Formula (Metric)

$$Q_D = 0.0037 \frac{T_b}{P_B}\left[\left(\frac{P_1^2 - P_2^2}{GTLZ}\right) d^{16/3}\right]^{1/2}$$

where:

Q_D = flow in m_3/day at Sc condition.

Panhandle Formula (Metric)

$$Q_D = 0.010E\left(\frac{T_b}{P_b}\right)^{1.02}\left(\frac{P_1^2 - P_2^2}{G^{0.961} LTZ}\right)^{0.510} d^{2.53}$$

Note: The symbols are same but units are in the standard metric units.

14.2.4 Two Phase Flow

Often the fluid flow to the plant is two phase. In this case the method of Flanigan[1] can be used. In this method the total pressure drop can be found from:

Equation 14-4

$$\Delta P_{TOTAL} = \Delta P_{TP} + \Delta P_h$$

where:

ΔP total = the total pressure drop.

ΔP_{TP} = the two phase pressure drop.

ΔP_h = the pressure drop due to changes in elevation over the length of the line.

The procedure is as follows:

For ΔP_{TP}

1. Find superficial gas velocity with no liquid present.
2. Find R - the BBLS of liquid in gas in BBLS/MMSCF.
3. Calculate $\frac{U}{R^{0.32}}$.
4. Read off pipeline efficiency from Figure 14.3
5. Calculate the pressure drop from the Panhandle Formula by inserting the proper pipeline efficiency.

For ΔP_h

Find from the formula:

Equation 14-5

$$\Delta P_h = \sum \frac{hF_e\rho_C}{144}$$

where:

h = the height of each separate hill.

F_e = the elevation correction factor obtained from Figure 14.5.

ρ_C = liquid density in lbs/cu.ft.

The metric version of the Flanigan method is as Follows:

For P_{TP}:

1. Find superficial gas velocity with no liquid present. $U = 5.18\frac{QTZ}{Pd^2}$, m/s, all terms as previously defined.
2. Find R, the liquid/gas ratio in aL/m^3.
3. Calculate $\frac{U}{R^{0.32}}$.
4. Read off pipeline efficiency from Figure 14.4.
5. Calculate the pressure drop from the Panhandle Formula by inserting the proper pipeline efficiency.

For ΔP_h:

Find from the formula:

$$\Delta P_h = \sum 0.009807\ hF_e\rho$$

where:

h = the height of each separate hill in meters

Fe = the elevation correction factor obtained from Figure 14.6.

ρ = the liquid density in kg/m^3

Note: Gregory and Aziz[2] have developed a method of calculating the pressure drop for pipelines based on the selection of the flow pattern and then using a formula applicable to the appropriate flow pattern, the pressure drop is calculated. This method is discussed in more detail in Chapter 6.

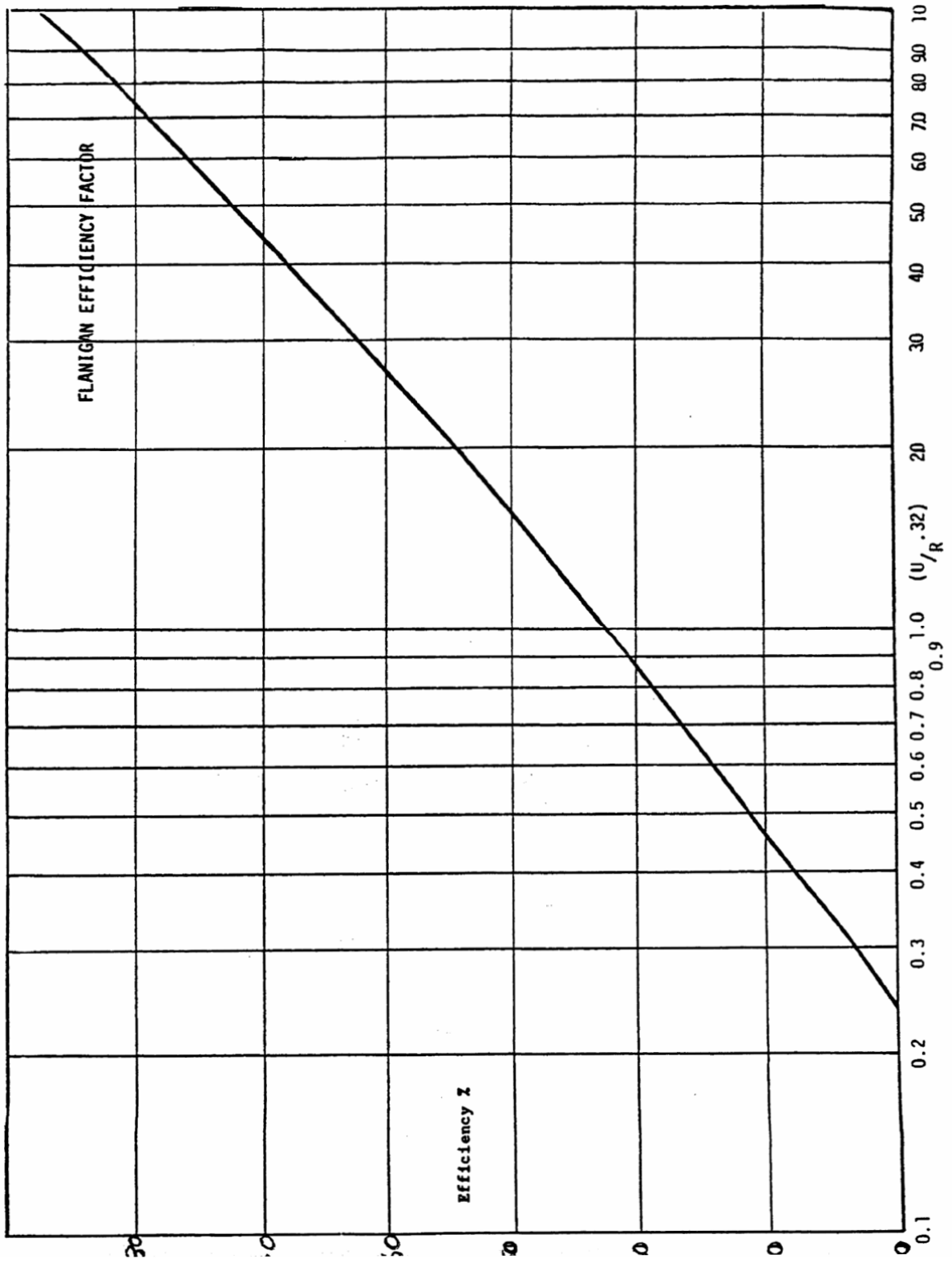


Figure 14.3: Flanigan Efficiency Factor

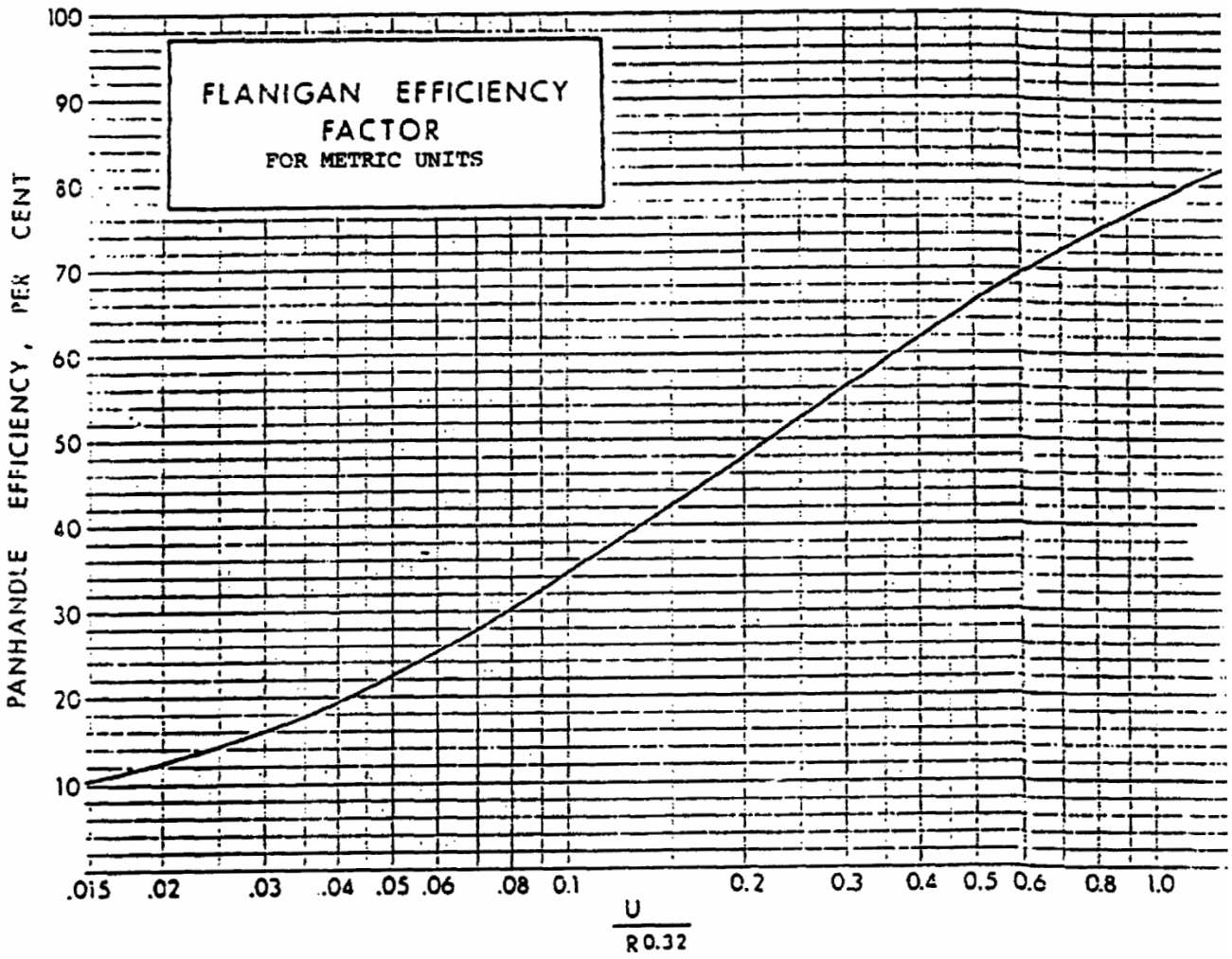


Figure 14.4: Flanigan Efficiency Factor (Metric)

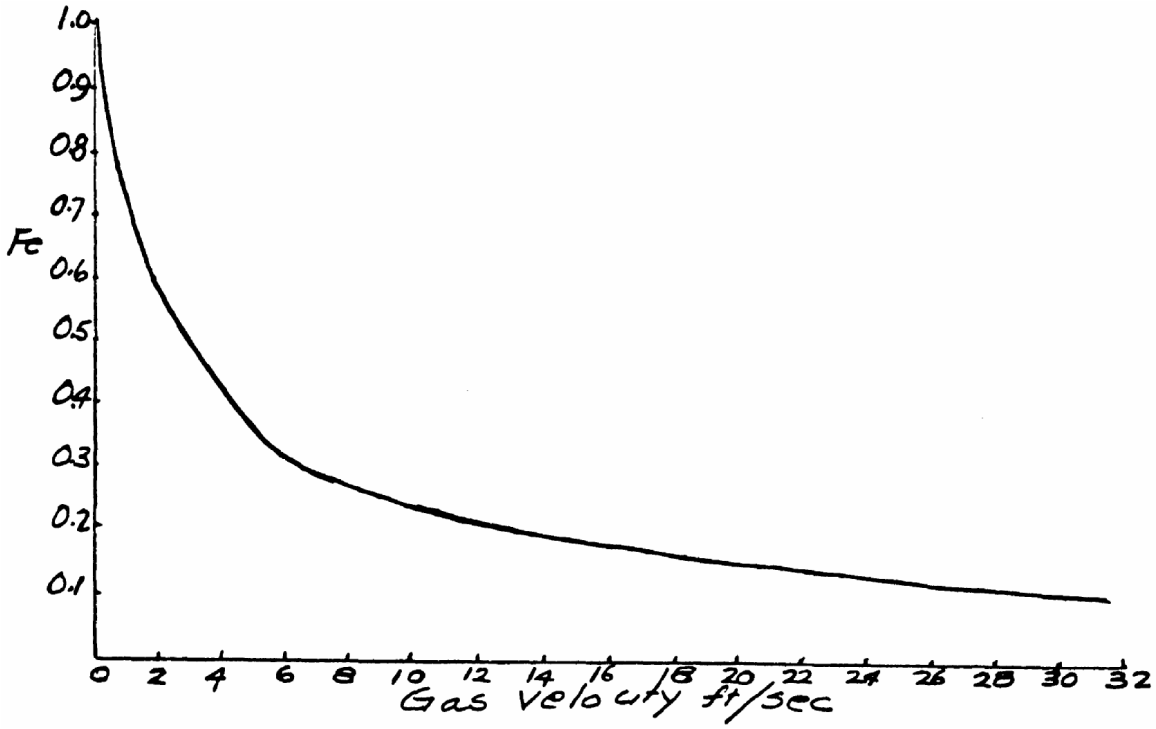


Figure 14.5: Elevation Correction Factor

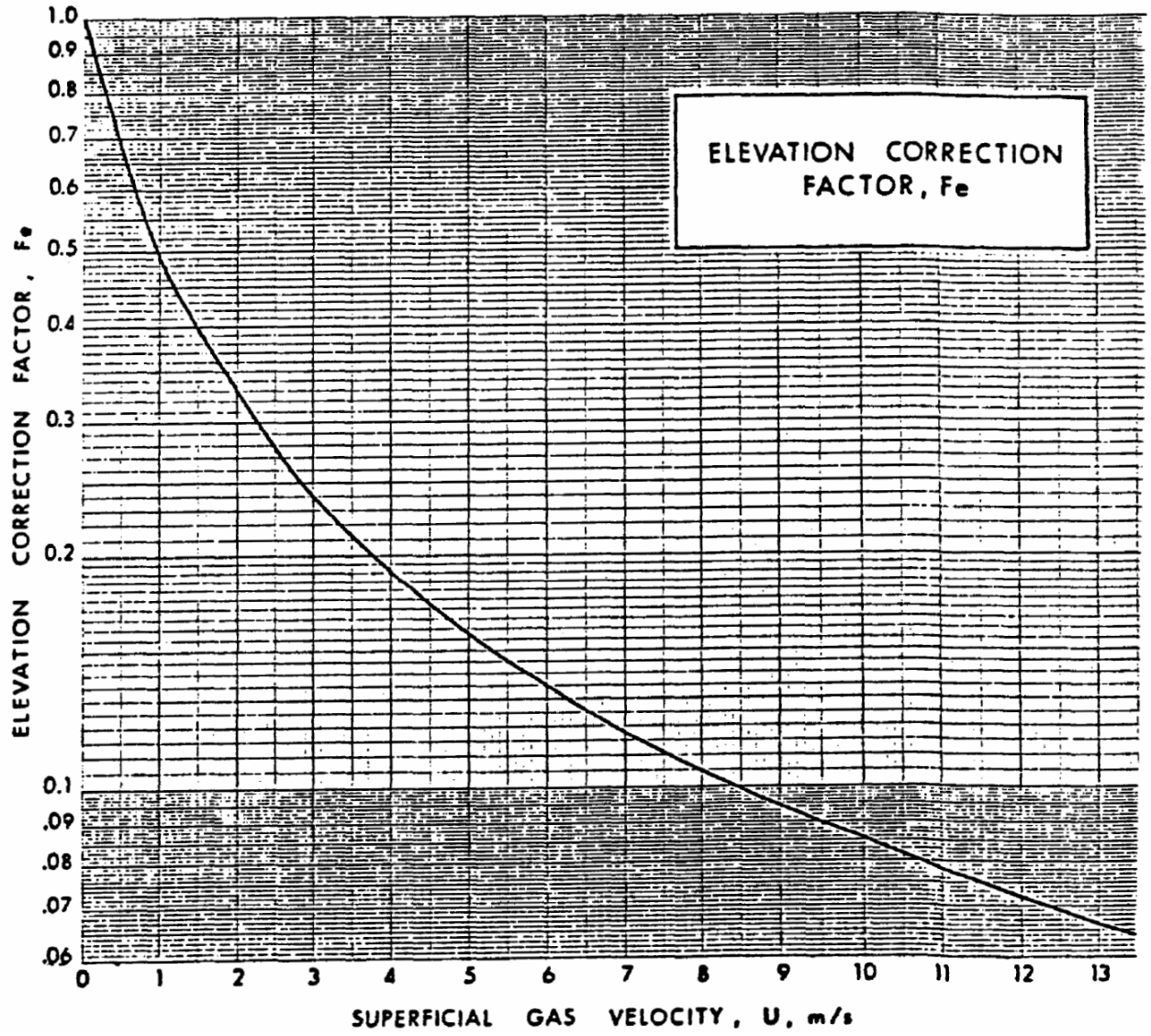


Figure 14.6: Elevation Correction Factor (Metric)

Example Problem 14.2

Size the lines given in Figure 14.1 with the design flow rates as worked out there. The gas is to be delivered to the plant at pressure of 1,000 psia. The area around the plant is relatively flat so that the effect in change of elevation will be neglected.

Solution

We will use the method of Flanigan. First, let us find the quantity of liquid that will probably be present. Run a flash at 1100 psia. This will give the BBLS of liquid per MMSCF. A flash of the fluid at 1100 psia and 700ºF gives the following data:

BBLS/MMSCF liquid in line	26.5
Liquid density (lb/cu.ft.)	33.83
The Z factor can be calculated and is	.85
Gas density (lb/cu.ft.)	4.94
The specific gravity of the gas	.734

From this data the pressure per mile for various lines, sizes and flows is calculated by means of a computer. The results of these calculations are given below.

Diameter (inches)	Q (MMCF/D)	R (B/MMCF)	Pressure Drop (PSI/Mile)
2.067	2.50	26.50	37.61
2.067	4.560	26.50	112.11
3.068	4.000	26.50	17.06
3.068	5.500	26.50	27.50
3.068	6.000	26.50	32.04
3.068	6.500	26.50	36.91
4.026	6.000	26.50	10.46
4.026	10.000	26.50	22.88
4.026	14.500	26.50	42.80
6.065	10.000	26.50	4.31
6.065	14.500	26.50	7.54
6.065	16.500	26.50	9.15
7.981	14.500	26.50	2.55
7.981	16.500	26.50	3.04

The pressure drop for various flows now given can be found by interpolation. Now working backwards from a plant entrance:

Line Description	Design Flow (MMSCFD)	Line Length (miles)	Recommended Line Size (inches)	Pressure Drop (psi)	Pressure at Wellsite (psia)
A-B	3.5	1.0	2"	75	--
B-D	2.0	1.0	2"	30	1105 psia
B-C	1.5	1.0	2"	20	1095 psia
A-G	14.5	1.0	4"	43	--
G-J	4.5	1.0	3”	20	1063 psia
G-H	10.0	1.0	4"	23	--

H-I	6.0	1.0	3”	37	1103 psia
A-K	5.5	0.5	3"	15	1015 psia
A-E	16.5	1.0	4"	58	--
E-F	6.5	1.5	3”	54	1112 psia

** Possibly could make this 2" in size.

14.3 Codes

To determine the thickness of pipe to use and for other general design requirements it is necessary to follow the appropriate code. This is the Canadian Standards Code GSA Z 184. This code now supersedes the previously used code in Canada which was the ANSI-B-31.8 - Code for Gas Transmission and Distribution piping. The Z 184 is very similar to the ANSI-B-31.8 code.

This code is not nearly as stringent as the codes used for plant piping which are:

ANSI-B-31 .3 Refinery Piping.

ANSI-B-31.1 Power Plant Piping.

The important part of the Z 184 Code† is the setting up of a system of design factors for different locations. The less dense the population or habitation, the higher allowable design yield. The classification of location for the pipe is given in Table 14.1. For hydrocarbon liquids, GSA Z 183† is used except for high vapour pressure liquids when often GSA Z 184 is used with special requirements for block or isolating valves.

14.4 Typical Pipeline Project

There are many facets from the initial concept of a pipeline project to its completion and initial operation. Figure 14.5 shows an arrow diagram of some of the more important divisions of the work. We will consider some of these in the development of a project.

14.4.1 Design

The pipe diameter is chosen as discussed previously by determining the allowable pressure drop. The pipeline code sets the allowable yield for the pipe to be used. However, there are many grades and types of pipe; generally either seamless or electric resistance welded pipe is used in gas gathering systems. There are also varying grades of pipe that can be used. Some of the more common ones are given in Table 14.3.

Table 14.1: Class Locations

General
The Class location unit is an area that extends 200 m on either side of the centreline of any continuous 1.6 km length of pipeline. The class location is determined by the buildings or dwelling units in the class location unit. Each dwelling unit in a multiple dwelling unit building is counted as a

† A new edition of this code has been published in 1992.

† A new edition of this code has been published in 1992.

separate unit.

Class 1

A Class 1 Location is:

a) Any class location that has ten or fewer dwelling units intended for human occupancy.

b) Any offshore pipeline.

Class 2

A Class 2 Location is:

a) Any class location unit that has more than 10 but fewer than 46 dwelling units intended for human occupancy; or

b) An area where the pipeline lies within 90 m of any of the following:

   a) A building that is occupied by 20 or more persons during normal use;

   b) A small, well defined outside area that is occupied by 20 or more persons during normal use, such as a playground, recreation area, outdoor theatre, or other places of public assembly.

Class 3

A Class 3 Location is any class location unit that has 46 or more dwelling units intended for human occupancy.

Class 4

A Class 4 Location is any class location unit where buildings with four or more storeys above ground are prevalent.

Adjustment of Boundaries

The boundaries of the class locations may be adjusted as follows:

a) A Class 4 Location ends 200 m from the nearest building with four or more storeys above ground.

b) When a cluster of buildings intended for human occupancy requires a Class 3 location, the Class 3 Location ends 200 m from the nearest building in the cluster, and

c) When a cluster of buildings intended for human occupancy requires a Class 2 Location, the Class 2 Location ends 200 m from the nearest building in the cluster.

Note: It is not intended here that a full mile of lower-stress-level pipeline shall be installed if there are physical barriers or other factors that will limit the further expansion of the more densely populated area to a total distance of less than 1 mile. It is intended, however, that where no such barriers exist, ample allowance shall be made in determining the limits of the lower-stress design to provide for probable further development in the area.

Note: Multistory means four or more "floors" above ground including the first or ground floor. The depth of basements or number of basement floors is immaterial.

The design factor and the classification type of construction is given in Table 14.2.

Table 14.2: Design Factors for Steel Pipe

Class Location	Design Factor (F)
1	0.72
2	0.60
3	0.50
4	0.40

A design factor of 0.60 or less shall be used in the design formula in Class 1 Locations that:

a) Crosses the right-of-way of an unimproved public road, without a casing.

b) Crosses without a casing, or has a parallel alignment on the right-of-way of either a hard surfaced road, a highway, or a public street.

c) Is supported by a vehicular, pedestrian, railway, or pipeline bridge; or

d) Is used in a fabricated assembly (including separators, pipeline valve assemblies, cross-connections, and river crossing headers; or is used within five pipe diameters in any direction from the last fitting of a fabricated assembly, other than a transition piece or an elbow used in place of a pipe bend which is not an associated assembly.

e) As a maximum for sour gas.

For Class 2 Locations, a design factor of 0.50, or less, shall be used in the design formula for uncased steel pipe that crosses the right-of-way of a hard surfaced road, a highway, or a public street.

For Class 1 or other Class Locations, a design factor of 0.50, or less, shall be used in the design formula for:

a) Compressor station, regulator station and measuring station gas piping, including the piping connecting the station to valving employed in isolating the station from the pipeline.

b) Offshore platform gas piping, including the pipe riser.

Note: For natural gas liquids there are no Class 2 locations if Class 1 does not apply the next one that has to be used is at least Class 3, ie. a design factor of 0.5 or lower.

Table 14.3: Common Pipe Classification

Canadian Designation	American Designation	Allowable Minimum Yield Strength PSI	(MPA)
CSA Z-245.3 Grade 207 Category I	API 5L Grade A	30,000	(207)
	ASTM A-53 Grade A	30,000	(207)
	ASTM A-106 Grade A	30,000	(207)
CSA Z-245.3 Grade 241 Category I	API 5L Grade B	35,000	(241)
	ASTM A-53 Grade B	35,000	(241)
	ASTM A-106 Grade B	35,000	(241)
CSA Z-245.3 Grade 290 Category I	API 5L x 42	42,000	(290)
CSA Z-245.3 Grade 317 Category I	API 5L x 46	46,000	(317)

CSA Z-245.3 Grade 359 Category I	API 5L x 52	52,000	(359)
CSA Z-245.3 Grade 414 Category I	API 5L x 60	60,000	(414)
CSA Z-245.3 Grade 207 Category II	ASTM A-333 Grade 1	30,000	(207)
CSA Z-245.3 Grade 241 Category II	ASTM A-333 Grade 6	35,000	(241)

The selection of the appropriate type of pipe is generally an economic one. However, a special review of all the conditions should be made before using high allowable stress pipe in sour gas service.

The design thickness is calculated from the following formula:

It is based on bursting force = containability

ie. *Pressure x Diameter = Steel Strength x Wall Thickness x 2*

Equation 14-6

$$P = \frac{2StFET}{d}$$

where:

P = design pressure in psia.

S = specified allowable minimum yield strength psia.

t = the design allowable wall thickness.

F = the construction factor design condition from the pipeline code.

T = the temperature derating factor, use 1.0 to 250ºF then derate.

E = the longitudinal join factor (1.00 for both seamless or ERW pipe and 0.80 for buttweld pipe).

d = diameter in inches.

Note: Data for pipe thickness is given on Fig. 10.16 GPSA Data Book p. 10-13, 16.

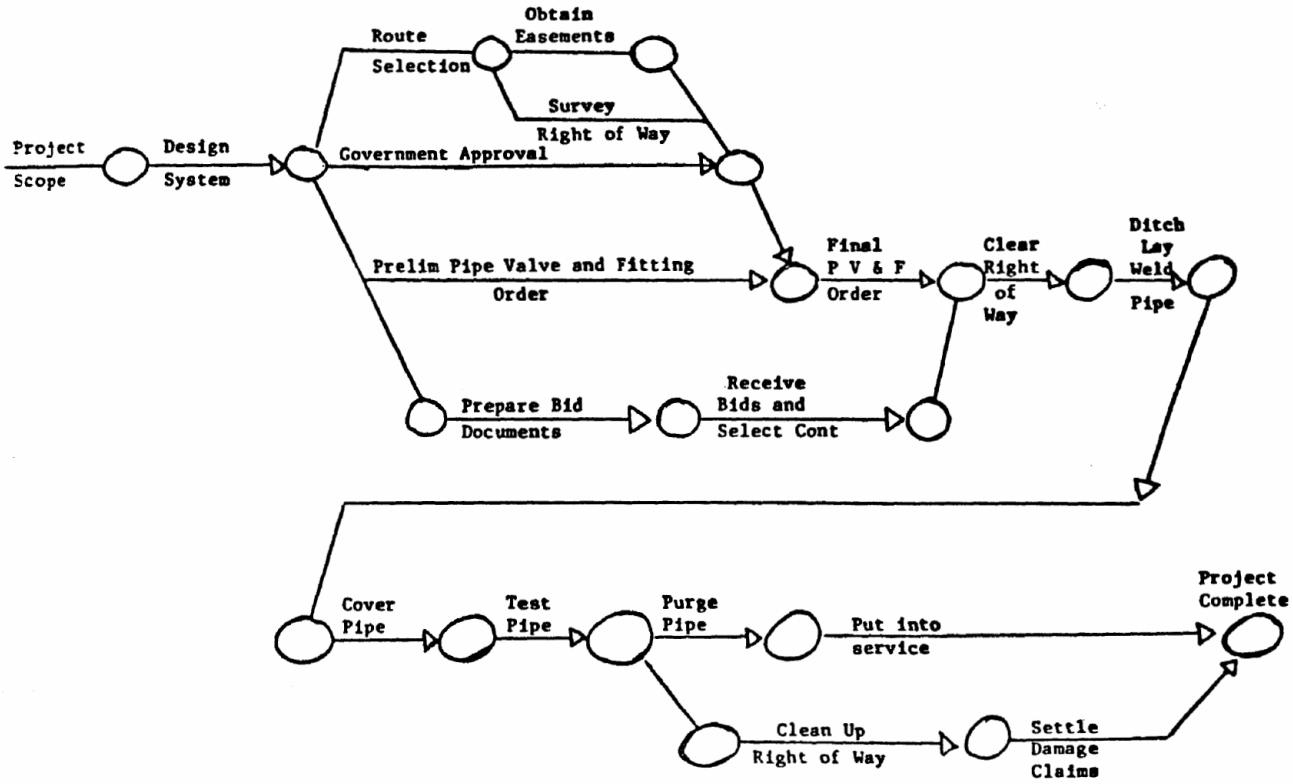


Figure 14.7: Typical Pipeline Project

When calculating the design thickness, the actual thickness used is the next thickest one according to the standard manufactured. One special note for sour gas service, generally consider adding a 1/8 corrosion allowance to the calculated pipe thickness before determining the design thickness.

In metric the design thickness is calculated from the following formula:

Equation 14-7

$$P = \frac{2St}{D} \times 10^3 \, FET$$

where:

P = design absolute pressure in kPa.

D = outside diameter of pipe in mm.

S = specified allowable minimum yield strength in MPa.

t = the design allowable wall thickness in mm.

F = the construction factor design condition from the CSA Z184 code.

E = Longitudinal joint factor obtained from the CSA Z184 code.

T = Temperature derating factor obtained from the CSA Z184 code.

When calculating the design thickness, the actual thickness used is the next thickest one according to the standard manufactured. One special note for sour gas service, generally consider adding a 2-3 mm corrosion allowance to the calculated pipe thickness before determining the design thickness.

Another item of interest is pressure rating of flanges. These are given in the Table 14.4 below.

Table 14.4: ANSI Rating VS M.0.P

Ansi Rating	Maximum Operating Pressure PSI/KPa	Test Pressure PSI/KPa
150	X1.83 = 275 X12.64 = 1896 KPA	X4.0 = 600 X27.58 = 4137 KPA
300	X2.4 = 720 X16.548 = 4964 KPA	X4.0 = 1200 X27.58 = 8274 KPA
600	X2.4 = 1440 X16.548 = 9929 KPA	X5.0 = 3000 X34.475 = 20,685 KPA
900	X2.4 = 2160 X16.548 = 14,893 KPA	X4.444 = 4000 X30.644 = 27,580 KPA
1500	X2.4 = 3600 X16.548 = 24,822 KPA	X4.0 = 6000 X27.58 = 41,370 KPA
2500	X2.4 = 6000 X16.548 = 41,370 KPA	X4.0 = 10,000 X27.58 = 68,950 KPA

14.4.2 Protection of Pipe

Normally steel pipe is externally coated for external corrosion protection. Some of the more common external coatings are:

- Coal tar.
- Polyethylene type tapes, either field or factory applied such as yellow jacket.

Normally, as an added means of protection, cathodic protection facilities are planned. These systems, of course, have to be custom designed to fit the soil conditions.

For internal corrosion protection special metallurgy is usually specified (for example as sour gas). The general specification is to try to obtain a soft steel that does not have high stresses in it. This is often met by specifying the following:

- The Rockwell C hardness should be below 20.
- Brinell (BHN) 225.
- Maximum yield strength 80,000 psi (very general).
- Meets NACE specs MR01-75 resistance to sulfide stress cracking.

The Carbon Equivalent should not exceed 0.45 and shall be determined by the following formula.

Equation 14-8

$$CE = C + \frac{Mn}{6} + \frac{Cr + Mo + Va}{5} + \frac{Ni + Cu}{15}$$

The selection of the appropriate type of pipe is generally an economic one. However, a special review of all the conditions should be made before using high allowable stress pipe in sour gas service. The specification for sour gas pipeline materials is covered by CSA Z 184, CSA 245.3 and CSA 245.5 .

14.4.3 Route Selection

The selection of a route for the pipe affects materially the costs of the pipeline. Preliminary selections are made by a detailed check of aerial photographs and topography maps. Particular care should be taken to avoid, if possible, muskegs, sloughs, steep banks, ravines, etc. After a preliminary route has been selected, the route selected should be inspected and any particular trouble spots checked. Generally a straight line type of layout is considered best. Whenever any length of line is to be installed, a helicopter is the best way to check the route. Also, soil conditions should be checked to determine the existence of rock, etc., in the ground. Special crossing header and valve location should also be reviewed.

14.4.4 Pipe. Valve and Fitting Ordering

After the design has been generally completed, the pipe should be placed on order. This is done by taking off from a scale map the amount of pipe required and adding about 3% for changes in grades, (this, of course, can vary with topography) and about 1% for wastage. If time permits, the survey should be done and the pipe ordered after surveying. This, however, is not often possible.

As soon as all the details for connection valving, etc. are complete, these materials should be placed on order.

14.4.5 Survey and Right of Way Acquisitions

Early in the project, a land department of the pipeline company must go out and obtain rights of way. These rights of way allow the crossing of the owners land. The exact location is determined by the survey crew. This crew follows the design maps and determines the exact location of the pipeline route. The survey party stakes each side of the right of way on which the pipeline is to be built. The width of the right of way varies from about 33 ft. to 75 ft. or more (10m to 25m).

Generally 50 ft. (45m) or greater right of way is desirable for good pipelining. This allows for reasonable working room without going off the allowable right of way.

14.4.6 Pipeline Construction

The next phase in the project is to prepare bid documents for the installation of the pipeline. The major parts of a pipeline bid document are:

a) The Agreement - This is the legal section dealing with the relations of the parties.

b) General Conditions - This spells out how the parties to the contract shall handle the work and the responsibilities of each party.

c) Specifications - This describes the specific work, where it is, what is to be done and the technical specification for such things as welding, coating, ditching, hauling and storing the pipe. The tie-in details, back filling, pressure testing, cleanup, etc.

d) Bidding Details - The actual price per foot for each size pipe and for each detail is obtained.

The actual installation of the pipeline is a specialized field and will not be further discussed. The pipeline bidding document governs most of the details. The field inspectors settle items not covered in the document.

14.4.7 Pressure Testing

A pressure test of the system must be carried out according to the requirements of Governmental regulations. The requirements in the Province of Alberta for gas pipelines are:

Service	Test Pressure
Sweet gas	1.25 x design pressure
Sour gas	1.4 x design pressure

The length of the pressure test is 24 hours.

The preferable test medium is water. However, because of winter conditions, it is often necessary to either use an antifreeze or a different fluid such as condensate, sweet gas or air. These fluids require particular care in their use as there are many hazards in using a hydrocarbon fluid or air as the test medium. A recent technique of using hot water has been used quite successfully by Nova.

14.4.8 Putting the Pipeline into Service

The first thing to do after the pipeline has been tested is to displace the test media and start purging. A detailed purging procedure for the pipeline should be developed. Even if the pipeline is to be used for sour gas it is very desirable to purge with sweet gas. There is a book that covers many facets of purging, published by the American Gas Association. Here are some of the main points of purging:

- Consider the use of an inert gas or steam purge for some applications.
- Keep the velocity in the pipeline down around 5-6 ft/sec.
- Purge until at least a reading of 95% gas is obtained.
- Make sure all lines and connections are suitably purged.

- Do not start pressurizing up until purging is finished.

14.4.9 Pressuring Up

After purging, the pipeline should be pressured up, either with sweet gas or the operating fluid, as a further check for leaks, and then put into service. It should be noted that when sour gas is used many further leaks in flanges, etc., will be found. These have to be tightened up before proceeding to put the line into use.

14.5 PROBLEMS

Example Problem 14.3

How thick would you make 12" API 5L x 52 ERW pipe in a sour gas service going across a private sparsely travelled trail. It is assumed that casing is not required. The operating pressure is 1200 psia. The normal operating temperature is 80ºF.

Example Problem 14.4

A gas reservoir of approximately 3.2 x 10^{12} SCF recoverable raw gas reserves has been discovered. The reservoir depth is 10,000 ft below surface. The gas composition is as given in Figure 1. Nine wells have been drilled. Those inside the 100' contour have a design deliverability against 1440 psig of 40 MMSCFD and those in the 10’ to 100' contour have a 20 MMSCFD deliverability. It is expected that all wells drilled in these respective areas will have the same deliverability characteristics.

The gas has been sold to Trans Canada Pipelines which has a maximum day demand of 1 SCF for every 8600 SCF of sales gas that can be recovered from the reservoir.

The gas composition is:

Compound	Mole %
N_2	0.54
H_2S	3.46
CO_2	11.50
C_1	80.25
C_2	3.65
C_3	0.35
iC_4	0.08
nC_4	0.12
iC_5^+	0.05
total	**100.00**

Size the most economical gathering system. Express your answer in diameter inch miles.

References

1. O. Flanigan, Oil & Gas Journal, March 10, 1958, P. 132.
2. G. A. Gregory and K. Aziz, Journal of Canadian Petroleum Technology, July-Sept. 1975, P. 29.
3. Purge Principles and Practices, American Gas Association, New York, 1954.

Reading List

1. R.L. Moore, "How to Design and Build Sour Gas Gathering Systems", Pipeline and Gas Journal, December 1977, P. 28.

15.0 Receiving Gas at a Plant

- Liquid Separations.
- Design of equipment.
- Operating problems.

15.1 Introduction

When a gas enters a plant, nearly always there is a large vessel through which the gas passes. The function of these vessels is to remove any free liquid from the gas stream before it is further processed. Since most gas enters the plant as a two phase fluid, these vessels are of prime importance. For example, if an inlet separator does not function properly and hydrocarbon liquid is carried over to the sweetening section of a plant, extensive foaming of the sweetening agent will occur. Also, the hydrocarbon will have to be removed through the still which will cause upsets in the sulphur plant, if one exists.

15.2 Function of Inlet Separators

These vessels have two functions.

These are:

1. Separate the fluid received into gas and liquid.
2. To handle liquid slugs and prevent the receiving of them at a plant from upsetting the rest of the plant.

15.3 Particle Size

When separating liquid and solids from a gas, the size of particles is of considerable importance. Figure 15.1 shows the range of particle sizes and the types of particles often encountered. Generally, we are concerned in the gas processing industry with the separation of liquid droplets - generally 10 microns in size (i.e, 1/100 of a millimeter or larger).

Note: This Chapter revised to metric by E. Wichert.

15.4 Types of Separators

There are several types of separators.

These are:

1. The gravity separator, which depends only on gravity for separation.
2. The impingement-type separator, in which the particles strike a barrier and then, by gravity, complete the separation.
3. Centrifugal separators, in which centrifugal force is used to separate the liquids.

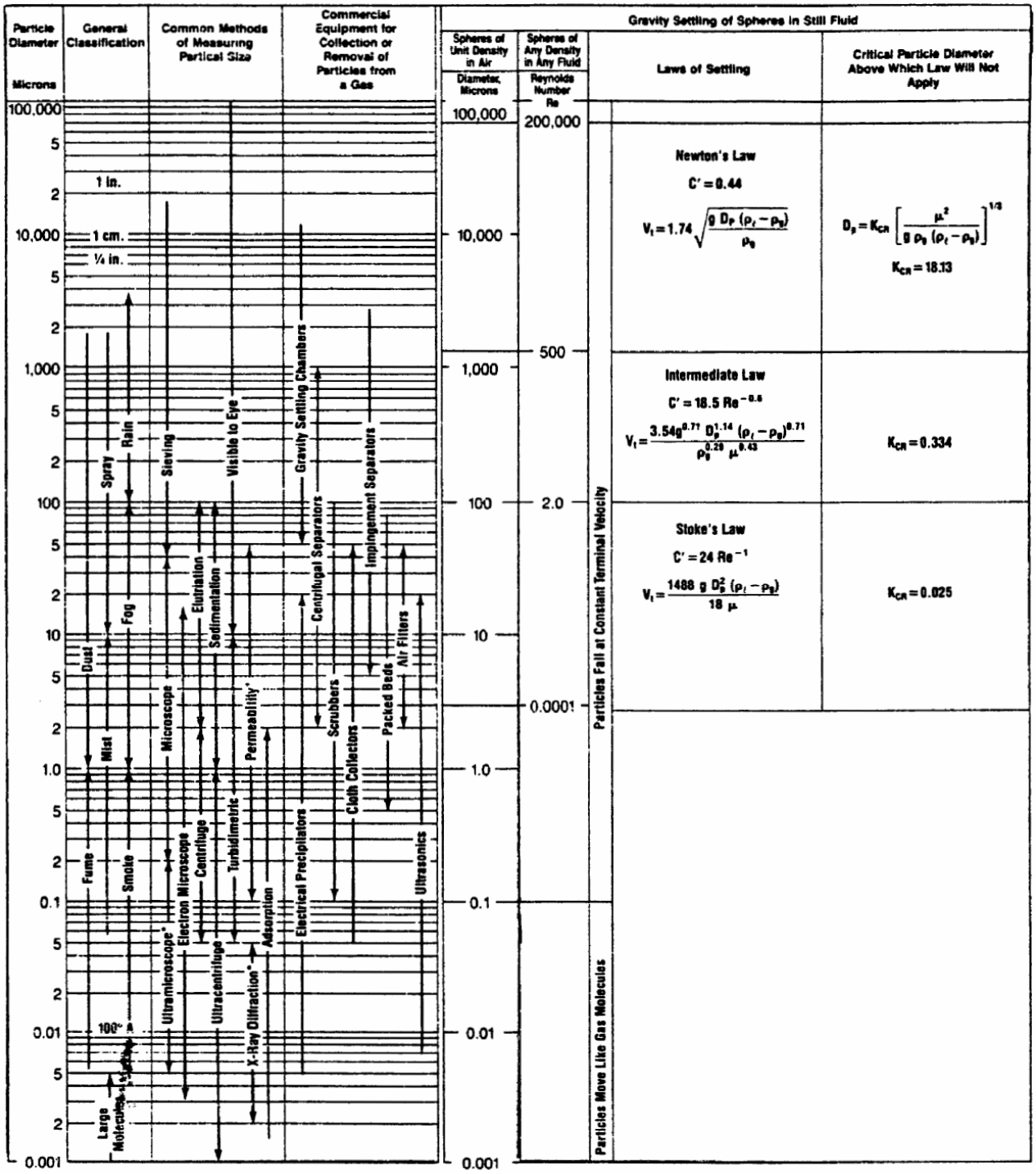


Figure 15.1: Gravity Settling Laws and Particle Characteristics

15.5 Description of Common Separators

In the gas processing industry generally some form of an impingement separator is used. Some of the more common ones are:

15.5.1 Vertical Separator

This is the most common type of separator and is normally used for the separation of gas from relatively large volumes of liquid. These vessels are normally installed at oil wells; not as much at gas wells. Also they are not good for handling slugs. The advantages of a vertical-type separator are that the liquid level control in the vessel is not too critical; it will handle large quantities of foreign material such as sand, without plugging up or otherwise affecting the operation of the vessel; and it is easy to clean. The disadvantages of a vertical separator are that it is the most expensive of the three types of separators; it does not adapt readily to a skid-mounted unit; and it has a lower capacity than any of the other kinds of separators when compared on the basis of effective diameter. Figure 15.3 illustrates the construction of a vertical separator.

The well effluent enters the vertical separator at approximately the midpoint of the vessel. Separation of the gas from the liquid commences at this point. The gas travels upwards through the vessel, dropping out the larger diameter liquid particles in its journey towards the top of the vessel. Most separators are designed on a basis of a ten micron diameter particle being excluded from the gas at exit point of the separator. All liquid droplets which are larger than ten microns will be either dropped out in the main body of the separator or will impinge upon the mist extractor and be removed at that point.

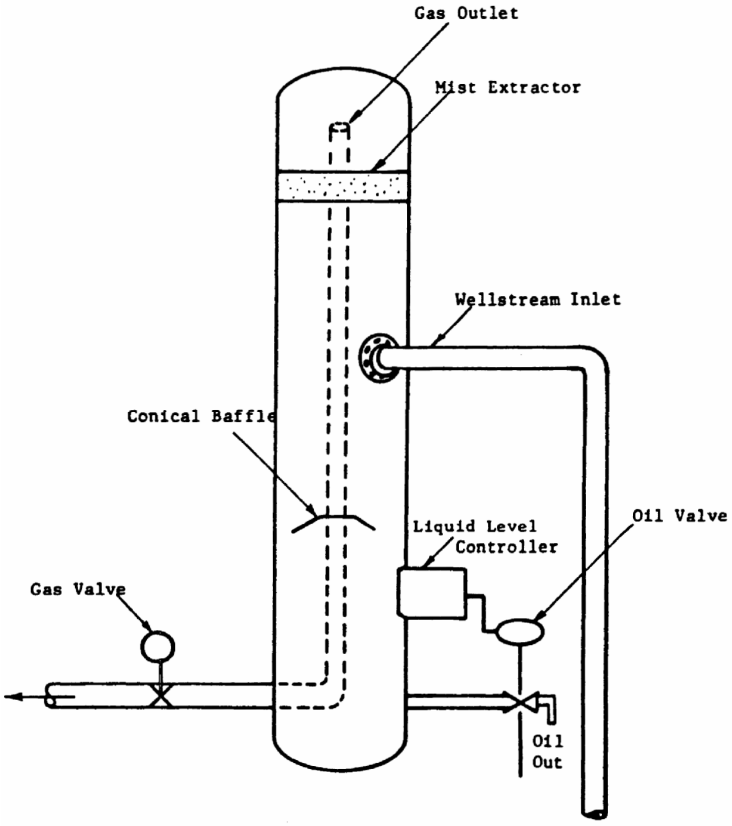


Figure 15.2: Vertical Separator

15.5.2 Horizontal Separator

The Horizontal separator is most commonly used for the separation of large volumes of gas from small volumes of liquid. It is also used extensively for handling liquid slugs from gathering systems.

This type of unit is used on gas wells, gas condensate wells, and, generally, as inlet separating devices at gas processing plants. The advantages of a horizontal unit are that it is cheaper than a vertical unit because it has approximately one and a half times the capacity for gas flow as that of a vertical separator of the same diameter; it is readily adaptable to skid mounting; and it is easier to hook up, due to the fact that the connections are more accessible.

The disadvantages of a horizontal separator are that the liquid level control is extremely critical compared to that of the vertical unit; and it is difficult to clean out once sand or other foreign material has accumulated in the vessel. Figure 15.3 illustrates the construction of a typical horizontal separator.

The well effluent enters one end of the horizontal separator, hits a deflector plate to drop out liquid drops, then the gas continues to the gas exit of the separator, where it passes through a mist extractor and then out the gas outlet. The liquid is collected in the bottom half and if necessary, a boot is put on at the outlet end to separate water from oil.

In the horizontal separator, the force of gravity acts on the liquid droplets throughout the length of the vessel causing the particles to "arc" to the bottom of the unit and thus separate more easily than in the vertical separator.

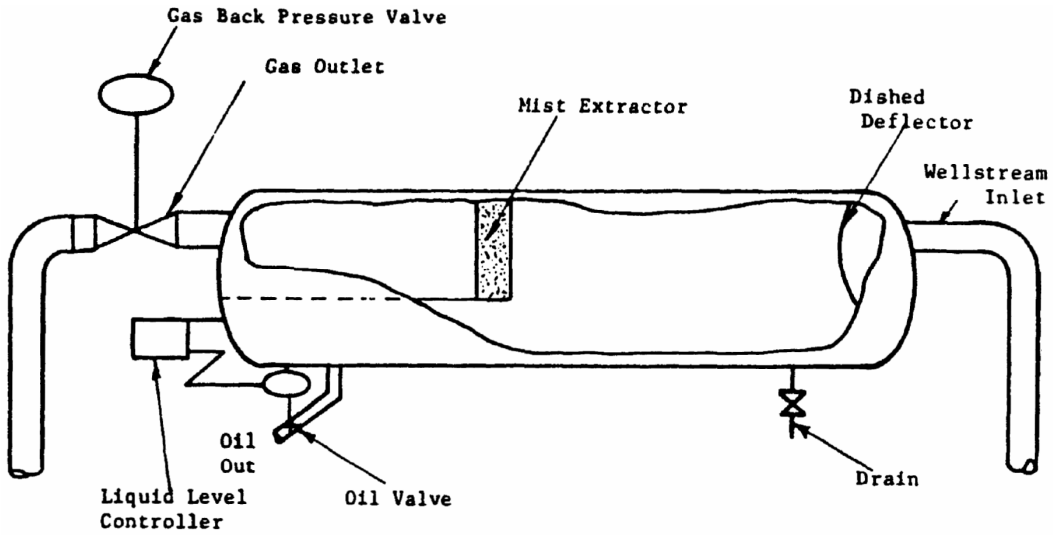


Figure 15.3: Horizontal Separator

15.5.3 Spherical Separator

The spherical separator is most commonly used for the separation of large volumes of gas from extremely small volumes of liquid. This type of unit is used mainly as a scrubber, i.e. removing small amounts of liquid. The advantage of a spherical separator is that it is more compact than either of the other units. It has the disadvantage of having a limited separation space and liquid surge capacity. In addition, the liquid level control is extremely critical. Figure 15.4 illustrates the construction of a typical spherical separator.

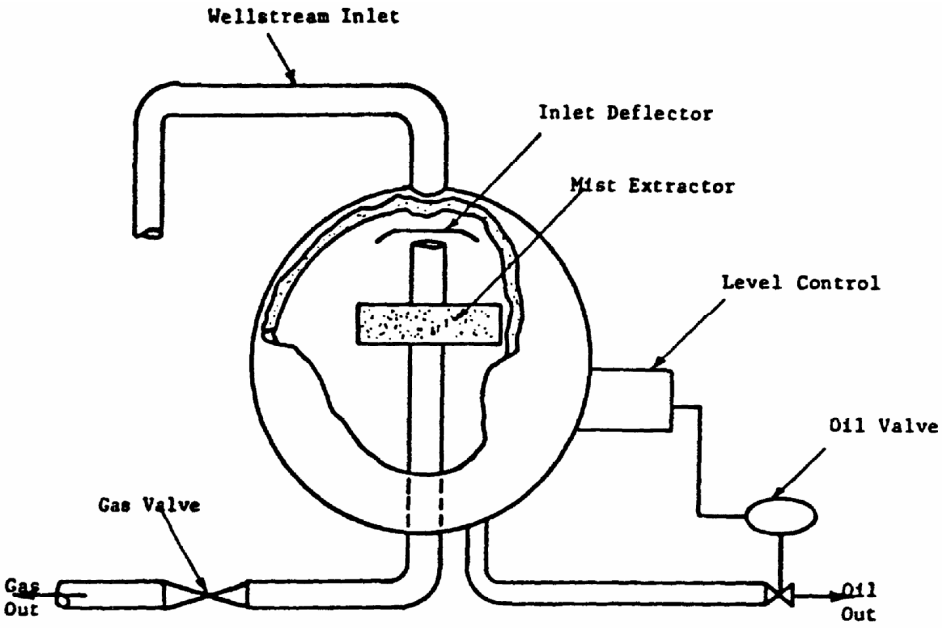


Figure 15.4: Spherical Separator

15.5.4 Cyclone Separator

The cyclone separator is a relatively new type of separating device as far as the gas industry is concerned. The principle of the cyclone has been utilized for years in other kinds of processing, however it has not been applied to gas processing until recently. It utilizes centrifugal force almost entirely to effect the separation between gas and liquid. This type of unit is used primarily as a scrubber, i.e., it is utilized for the separation of small volumes of liquid from relatively large volumes of gas. These units can be constructed very economically as it requires a relatively small diameter vessel to effect adequate separation. The relatively small size of these units makes them particularly adaptable to portable well test units. Figure 15.5 illustrates the construction of a cyclone type separator.

The well effluent enters the top of the vessel through a tangential inlet. This method of entry causes the effluent to be spun around the interior of the vessel. The centrifugal force exerted by this spinning causes the liquid to cling to the wall of the vessel. The liquid then drains by the force of gravity to the bottom of the vessel where it is drawn off by a liquid level operated dump valve. In place of a mist extractor which was used in the construction of the previous type of separators which were discussed, some of the cyclone separators utilize a recycling system. This operates as follows.

Near the gas outlet a slot is cut in the exit pipe. The pressure on the outside of the pipe, i.e. in the interior of the separator, is less than the pressure on the inside of the pipe. This is caused by a vortex which has been created due to the gas spinning within the "vortex chamber" of the vessel. The small liquid droplets which have remained in the gas stream are pulled out through the gap in the gas outlet pipe and recycled to the lower section of the separator where they drain by gravity down to the bottom of the vessel to join the main liquid stream. This particular method of mist extraction is a patented process, the patent being held by Portatest Manufacturing Ltd., Edmonton, Alberta. Figure 15.6 shows this action.

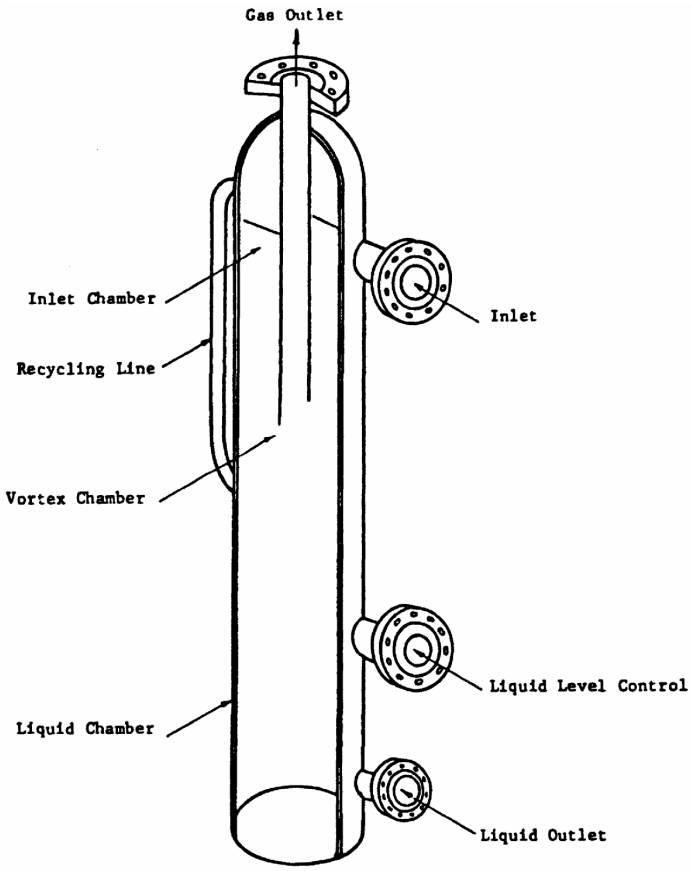


Figure 15.5: Cyclone Separator

15.6 Separation Theory

15.6.1 Momentum

Fluid phases with different densities will have a different momentum. If a two phase stream changes direction sharply greater momentum will not allow the particles of the heavier phase to turn as rapidly as the lighter fluid, so separation occurs. Momentum is usually employed for bulk separation of the two phases in a stream.

15.6.2 Gravity Settling

Liquid droplets will settle out of a gas phase if the gravitational force acting on the droplet is greater than the drag force of the gas flowing around the droplet (see Figure 15.6).

These forces can be described mathematically using the terminal or free settling velocity.

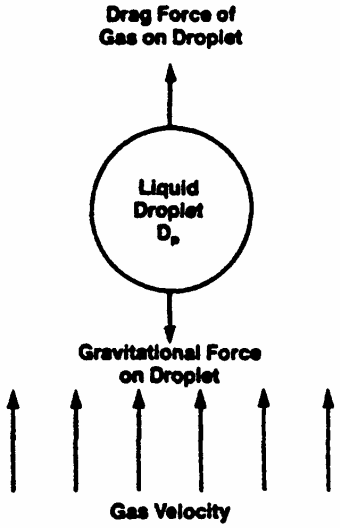


Figure 15.6: Forces on Liquid Droplet in Gas Stream

Equation 15-1

$$V_g = \sqrt{\frac{2g\,M_p\left(\rho_l - \rho_g\right)}{\rho_l \rho_g A_p C'}} = \sqrt{\frac{4gD_p\left(\rho_l - \rho_g\right)}{3\rho_g C'}}$$

The drag coefficient has been found to be a function of the shape of the particle and the Reynolds number of the flowing gas. For the purpose of this equation particle shape is considered to be a solid, rigid sphere.

Reynolds number is defined as:

Equation 15-2

$$\mathrm{Re} = \frac{1488\,D_p V_t \rho_g}{\mu}$$

In this form, a trial and error solution is required since both particle size, D_p, and terminal velocity, V_t, are involved. To avoid trial and error, values of the drag coefficient are presented in Figure 15.7 as a function of the product of drag coefficient, C', times the Reynolds number squared; this eliminates velocity from the expression: The abscissa of Figure 15.7 is given by:

Equation 15-3

$$C'\left(\mathrm{Re}\right)^2 = \frac{\left(0.95\right)\left(10^6\right)\rho_g D_p^3\left(\rho_l - \rho_g\right)}{\mu^2}$$

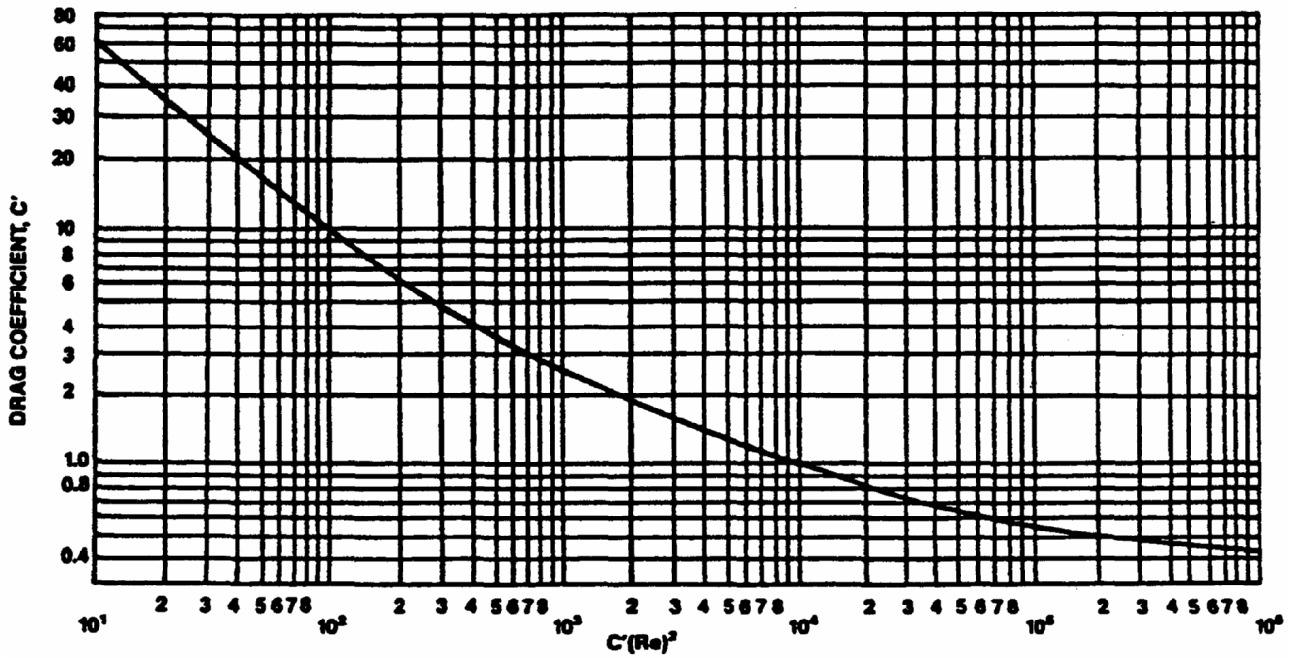


Figure 15.7: Drag Coefficient of Rigid Spheres

15.6.3 Gravity Settling - Limiting Conditions

As with other fluid flow phenomena, the drag coefficient reaches a limiting value at high Reynolds numbers.

15.6.3.1 Newton's Law

For relatively larger particles (approximately 1000 micron and larger) the gravity settling is described by Newton's law (see Figure 15.8). The limiting drag coefficient is 0.44 at Reynolds numbers above about 500. Substituting C'. = 0.44 in Eq. 15.1 produces the Newton's law equation expressed as:

Equation 15-4

$$V_t = 1.74 \frac{g D_p \left(\rho_l - \rho_g\right)}{\rho_g}$$

An upper limit to Newton's law is where the droplet size is so large that it requires a terminal velocity of such magnitude that excessive turbulence is created. The maximum droplet which can settle out can be determined by:

Equation 15-5

$$D_p = K_{CR}\left[\frac{\mu^2}{g\rho_g\left(\rho_l - \rho_g\right)}\right]^{1/3}$$

For the Newton's law region, the upper limit to Reynolds number is 200,000 and K_{CR} = 18.13 .

15.6.3.2 Stoke's Law

At low Reynolds numbers (less than 2), a linear relationship exists between drag coefficient and the Reynolds number (corresponding to laminar flow). Stoke's law applies in this case and Eq. 15.1 can be expressed as:

Equation 15-6

$$V_t = \frac{1488 g D_p^2 \left(\rho_l - \rho_g\right)}{18\mu}$$

The droplet diameter corresponding to a Reynolds number of 2 can be found using a value of 0.0080 for K_{CR} in Eq. 15.5.

The lower limit for Stoke's law applicability is a droplet diameter of approximately 3 microns. The upper limit is about 100 microns.

15.6.3.3 Coalescing

Very small droplets such as fog or mist cannot be separated practically by gravity. These droplets can be coalesced to form larger droplets that will settle by gravity. Coalescing devices in separators force gas to follow a tortuous path. The momentum of the droplets causes them to collide with other droplets or the coalescing device, forming larger droplets. These larger droplets can then settle out of the gas phase by gravity. Wire mesh screens, vane elements, and filter cartridges are typical examples of coalescing devices.

15.6.3.4 Separator Design and Construction

Separators are usually characterized as vertical, horizontal, or spherical. Horizontal separators can be single or double barrel and can be equipped with sumps or boots.

15.6.3.5 Parts of a Separator

Regardless of shape, separation vessels usually contain four major sections, plus the necessary controls. These sections are shown for horizontal and vertical vessels in Figure 15.6. The primary separation section, A, is used to separate the main portion of free liquid in the inlet stream. It contains the inlet nozzle which may be tangential, or a diverter baffle to take advantage of the inertial effects of centrifugal force or an abrupt change of direction to separate the major portion of the liquid from the gas stream.

The secondary or gravity section, B, is designed to utilize the force of gravity to enhance separation of entrained droplets. In some designs, straightening vanes are used to reduce turbulence. The vanes also act as droplet collectors, and reduce the distance a droplet must fall to be removed from the gas stream.

The coalescing section, C, utilizes a coalescer or mist extractor which can consist of a series of vanes, a knitted wire mesh pad, or cyclonic passages. This section removes the very small droplets of liquid from the gas by impingement on a surface where they coalesce. A typical liquid carryover from the mist extractor is less than 0.1 gallon per MMSCF.

The sump or liquid collection section, D, acts as receiver for all liquid removed from the gas in the primary, secondary, and coalescing section. Depending on requirements, the liquid section should have a certain amount of surge volume, for degassing or slug catching, over a minimum liquid level necessary for controls to function properly. Degassing may require a horizontal separator with a shallow liquid while emulsion separation may also require higher temperature, higher liquid level, and/or the addition of a surfactant.

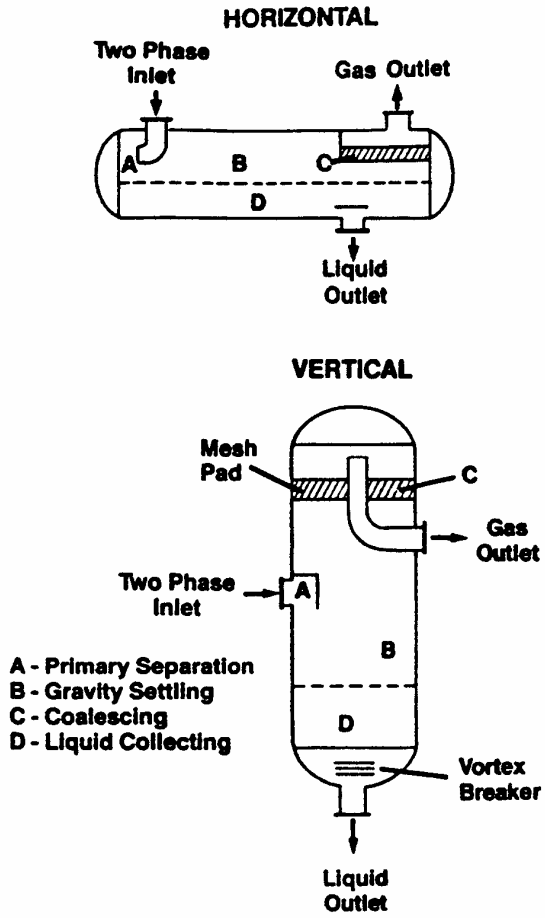


Figure 15.8: Gas-Liquid Separators

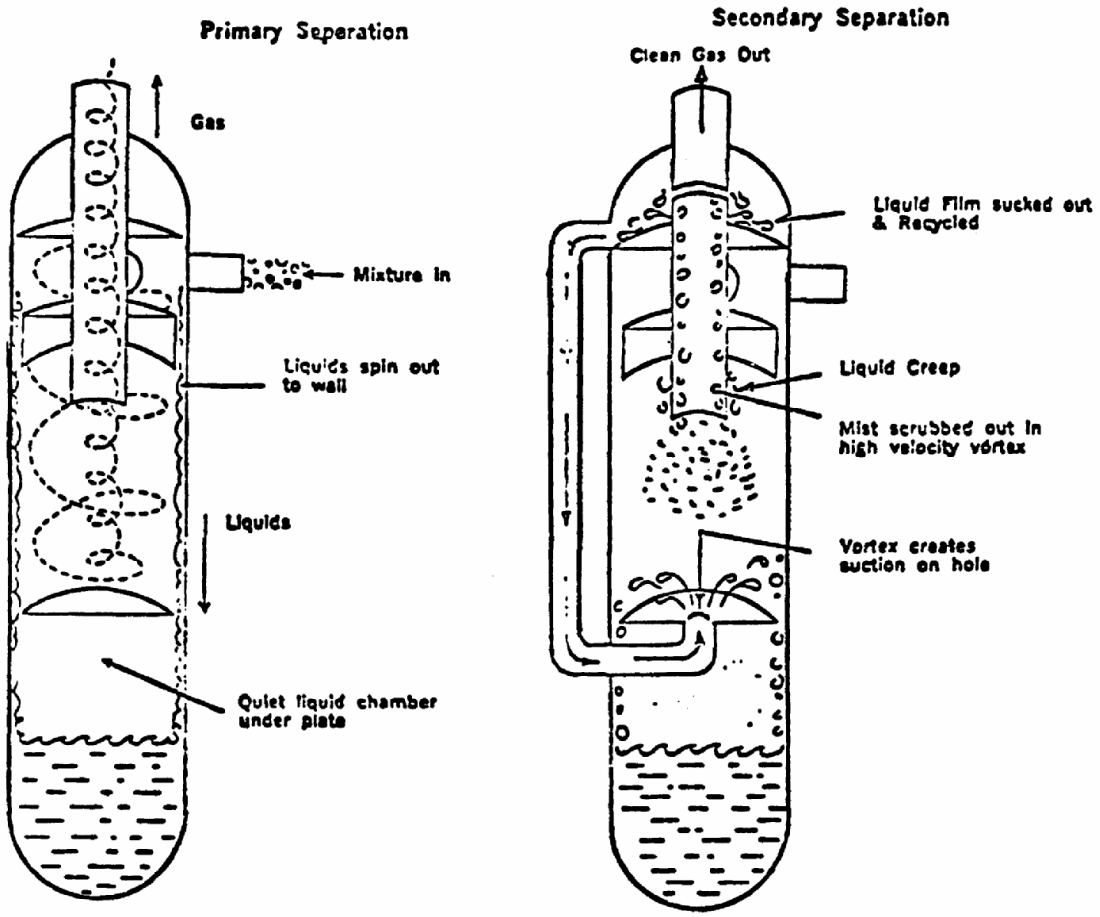


Figure 15.9: Recycling Action in Centrifugal Separator

15.7 Design of the Separator

There are two design criteria.

15.7.1 Allowable Velocity

The first is to determine the allowable velocity. This is done by means of a form of the Stokes' Law which can be expressed by the following equation:

Equation 15-7

$$u = K\sqrt{\frac{\rho_L - \rho_g}{\rho_g}}$$

where:

u = allowable velocity, feet per second or (m/s).

ρ_g = density of the gas, lb/cu ft. (kg/m^3).

ρ_L = density of the liquid, lb/cu ft. (kg/m^3).

K = a constant.

The common values for “K” are:

Vertical Separators without Mesh	
For a minimum of entrainment	0.10 (0.03)
Acceptable	0.175 Use either acceptable (0.053) or moderate for design (0.061)
Moderate	0.200 (0.061)
Appreciable	0.2500 (0.076)
Vertical Separators with Mesh (6" generally)	
Acceptable	0.35 (0.107)
Moderate	0.40 (0.122)
Horizontal Separators	
Acceptable	0.50 (0.152)

When a mesh is added to a horizontal vessel, "K" is not normally changed. The mesh generally acts as a secondary collection means.

Sometimes it is desirable to correlate the "K" factor with the flow of liquid and vapour. Watkins gives a graph for determining this.

By use of the previous equation and the actual flow, the area of the vessel can be determined from:

Equation 15-8

$$\frac{Q}{u} = A$$

where:

A = the cross sectional area of the vessel (m^2).

Q = the actual flow in cubic feet per second (m^3/s).

Note: For horizontal vessel only, 1/2 the area is for disengaging. The other half is used for surge. In vertical vessels allow at least 4 ft of disengaging space. Figure 15.10 and Figure 15.11 illustrate the points mentioned above.

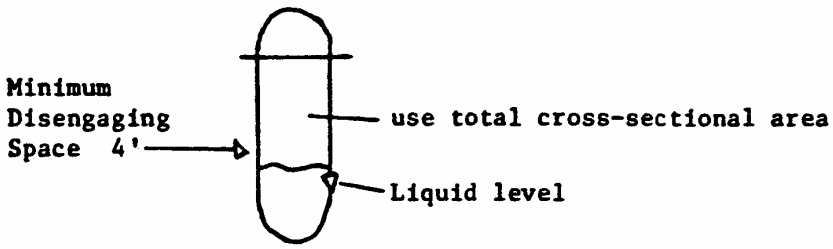


Figure 15.10: Disengaging Space Vertical Separation

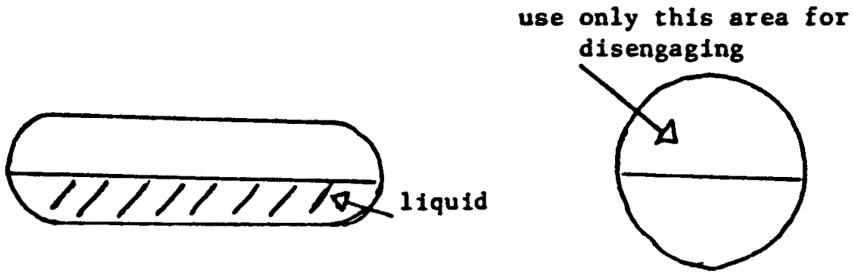


Figure 15.11: Using Half Disengaging Space Horizontal Separation

15.7.2 Vessel Height or Length

The second design criteria is to calculate the vessel height or length by determining the actual volume required for holding liquids. This can be set by:

- Providing a resident time of 1/2 hour so as to allow water and glycol to separate from the hydrocarbon condensate.
- Determining the amount of a slug that may be reasonably expected. This can be determined by the use of the Baker Liquid Hold Up Chart (Figure 15.9). This is done by calculating the amount of liquid that the line holds at a low flow rate and the amount of liquid that the line will hold at a high flow. The difference is the amount of liquid that will come into the separator in the period of time the flow rate is changed. Another method of calculating the hold up is by the method of Gregory and Aziz. A recent series of articles on liquid hold-up in pipelines has been published in the Oil and Gas Journal.

Example Problem 15.1

Size the horizontal inlet separator for a plant that receives raw natural gas that contains liquid from a 6 mile long 10" line. The average flow is 100 MMSCFD, the minimum is 70 MMSCFD and the maximum is 133 MMSCFD. The minimum time for going from minimum to maximum flow is one hour. The operating pressure is 1000 psia and the operating temperature is 70ºF.

The gas phase has a molecular weight of 20 and is sweet. The critical properties of the gas are 650 psia and 450ºF. The density of the liquid at actual operating conditions is 40 lbs/cu ft.

Solution

We have to determine the vessel's diameter and its length. To determine the diameter we use the allowable velocity formula and apply it over 1/2 the cross sectional area as we have a horizontal vessel:

$$u = K\sqrt{\frac{\rho_L - \rho_g}{\rho_g}}$$

Now

K = 0.50 for a horizontal vessel.

ρ_L = 40.

Must find P_g

First find the compressibility factor for the gas by means of reduced pressures and temperatures:

$$T_R = \frac{530}{450} = 1.18$$

$$P_R = \frac{1000}{650} = 1.54$$

Z = 0.63 (from P. 16-8 Fig. 16.3 NGPSA Manual).

$$\rho_g = \frac{(1000)(520)(20)}{(14.67)(379)(530)(0.63)} = 5.60\ lbs/cu\ \ ft.$$

$$u = 0.5\sqrt{\frac{40 - 5.6}{5.6}}$$

$$u = 1.25\ \ ft/\sec.$$

Now the actual flow at maximum rates is

$$\frac{133 \times 10^6}{(379)} \quad \left(\frac{20}{5.60}\right) = 1.26 \times 10^6\ \ ACFD$$

Area required for 1/2 vessel

$$\frac{1.26 \times 10^6}{(1440)(60)(1.25)} = 11.8\ sq.\ ft.$$

Therefore diameter = 5.5 ft.

Now for simplification we will assume that the pressure drop for gas and liquid separately in the gathering system is as follows (should calculate):

Gas Flow (MMSCFD)	ΔP_L if only liquid present (psi)	ΔP_g if only gas present (psi)
70	6.2	40
100	7.5	61
133	8.5	90

Now we shall use the Baker Liquid Hold Up Chart (Figure 15.12) and find:

$$X = \sqrt{\frac{\Delta P_L}{\Delta P_g}}$$

Find "X" at the two different flows

At 70 MMSCFD

$$X = \sqrt{\frac{6.2}{40}} = 0.39$$

At 133 MMSCFD

$$X = \sqrt{\frac{8.5}{90}} = 0.31$$

From the Hold Up Curve at 133 MMSCFD liquid will occupy 19% of the pipe line volume.

At 70 MMSCFD liquid will occupy 22% of the pipe line volume.

Therefore liquid that must come out during the change in flow:

0.03 (6) (520) = 93.4 barrels.

where:

520 bbls = the volume of one mile of ten inch pipeline.

Therefore the length of the vessel:

$$\frac{93.4\,(5.6)}{11.8} = 44.4\ ft$$

You could say 45 ft. long.

Thus, you should allow a vessel 5½ ft in diameter by 45 ft. long. The metric version of this problem is done below.

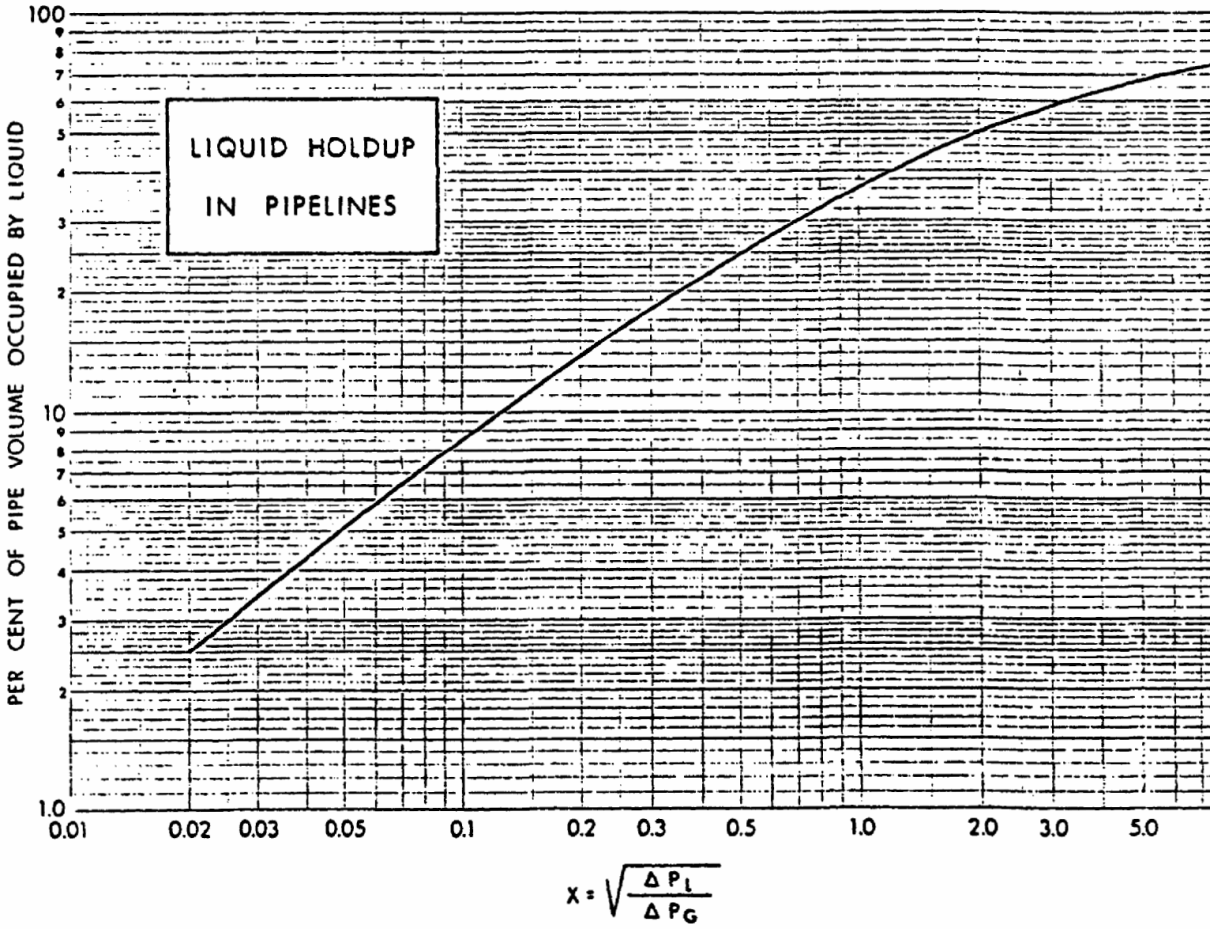


Figure 15.12: Liquid Hold-up in Pipelines

Example Problem 15.2

(Metric Version)

Size the horizontal inlet separator for a plant that receives raw natural gas that contains 300 mL of liquid per m^3 of gas from a 10 km long 254 mm diameter line. The average flow is 2800 10^3 m^3/d, the minimum is 2000 10^3 m^3/d and the maximum is 3750 10^3 m^3/d. The minimum time for going

from minimum to maximum flow in one hour. The operating pressure is 6900 kPa and the operating temperature is 231ºC.

The gas phase has a molecular mass of 20 and is sweet. The critical properties of the gas are 4482 kPa and 250 K. The density of the liquid at actual operating conditions is 641 kg/m^3.

Solution

We have to determine the vessel's diameter and its length. To determine the diameter we use the allowable velocity formula and apply it over 1/2 the cross sectional area as we have a horizontal vessel:

$$u = K\sqrt{\frac{\rho_L - \rho_g}{\rho_g}}$$

Now

K = 0.152 for a horizontal vessel.

ρ_L = 641 kg/m^3.

Must find P_g

First find the compressibility factor for the gas by means of reduced pressures and temperatures:

$$T_R = \frac{296.15}{250} = 1.18$$

$$P_R = \frac{6900}{4482} = 1.54$$

Z = 0.63 (from P. 16-8 Fig. 16.3 NGPSA Manual).

$$\rho_g = \frac{MP}{8.3145\,T\,Z} = \frac{20 \times 6900}{8.3145 \times 296.15 \times 0.63} = 89\ kg/m^3$$

$$u = 0.152\sqrt{\frac{641 - 89}{89}}$$

$$u = 0.378\ m/s$$

Now the actual flow at maximum rates is

$$Q_a = 0.00407\frac{QTZ}{P} \times \left(\frac{370 \times 296.15 \times 0.63}{6900}\right) = 0.413\ m^3/s$$

Area required for 1/2 vessel

$$\frac{0.413}{0.379} = 1.09\ m^2$$

Therefore diameter = 1.67 m or 1670 mm.

Now for simplification we will assume that the pressure drop for gas and liquid separately in the gathering system is as follows (should calculate):

Gas Flow (10^3m^3/d)	ΔP_L if only liquid present (KPa)	ΔP_g if only gas present (KPa)
2000	42.7	275.8

2800	51.7	420.6
3750	58.6	620.6

Now we shall use the Baker Liquid Hold Up Chart (Figure 15.12) and find:

$$X = \sqrt{\frac{\Delta P_L}{\Delta P_g}}$$

Find "X" at the two different flows

At 2000 $10^3 m^3/d$

$$X = \sqrt{\frac{42.7}{275.8}} = 0.39$$

At 133 MMSCFD

$$X = \sqrt{\frac{58.6}{620.6}} = 0.31$$

From the Hold Up Curve at 3750 $10^3 m^3/d$ liquid will occupy 19% of the pipe line volume.

At 2000 $10^3 m^3/d$ liquid will occupy 22% of the pipe line volume.

Volume of 10 km of 254 mm ID line is:

$$V = \frac{10000\pi}{4}\left(\frac{254}{1000}\right)^2 = 507\ m^3$$

Therefore liquid that must come out during the change in flow:

0.03 x 507 = 15.2 m^3.

Therefore the length of the vessel:

$$\frac{15.2}{1.09} = 13.9\ m$$

You could say 14 meters long.

15.8 Filter Separators

This type of separator has a higher separation efficiency than the centrifugal separator, but it uses filter elements, which must periodically be replaced. An example filter separator is shown in Figure 15.13. Gas enters the inlet nozzle and passes through the filter section where solid particles are filtered from the gas stream and liquid particles are coalesced into larger droplets. These droplets pass through the tube and are entrained into the second section of the separator; where a final mist extraction element removes these coalesced droplets from the gas stream.

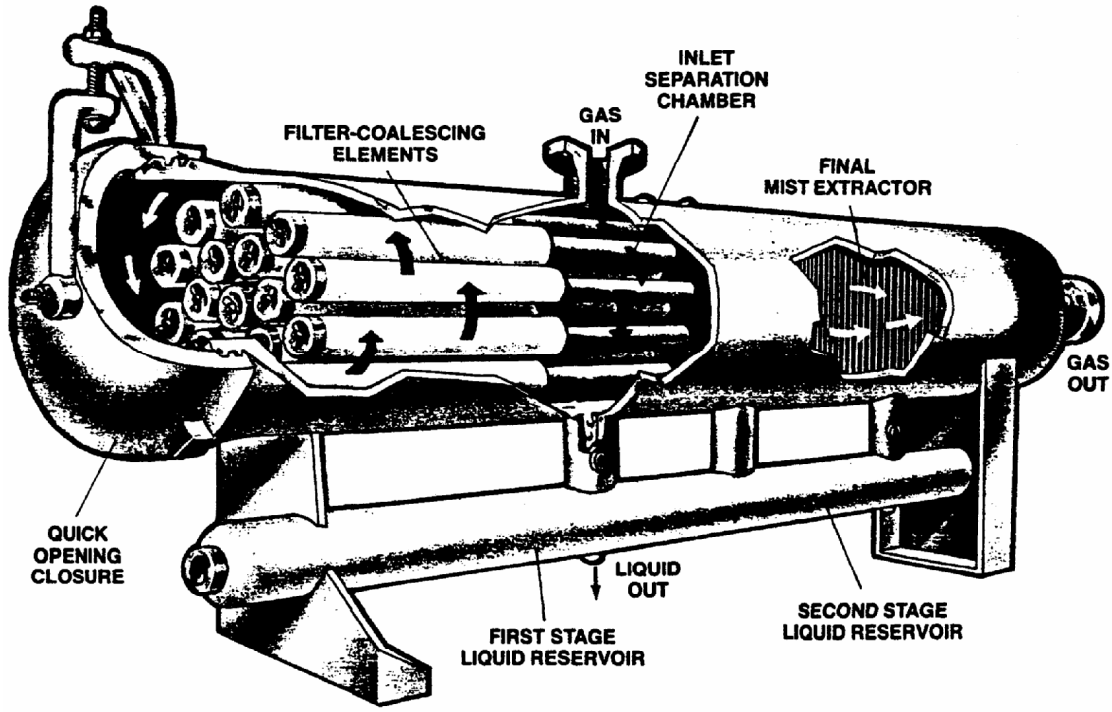


Figure 15.13: Example Horizontal Filter Separator

The design of filter separators is proprietary and a manufacturer should be consulted for specific size and recommendations. The body size of a horizontal filter separator for a typical application can be estimated by using 1.3 for the value of K in 15.6. This provides an approximate body diameter for a unit designed to remove water (other variables such as viscosity and surface tension enter into the actual size of determination). Units designed for water will be smaller than units sized to remove light hydrocarbons.

In many cases the vessel size will be determined by the filtration section rather than the mist extraction section. The filter cartridges coalesce the liquid mist into droplets which can be easily removed by the mist extractor section. A design consideration commonly overlooked is the velocity out of these filter tubes into the mist extraction section. If the velocity is too high, the droplets will be sheared back into a fine mist that will pass through the extractor element. A maximum allowable velocity for gas exiting the filter tube attachment pipe can be estimated using momentum equations hydrocarbon liquids or low pressure gas should be limited to even less than this value. No published data can be cited since this information is proprietary with each filter separator manufacturer.

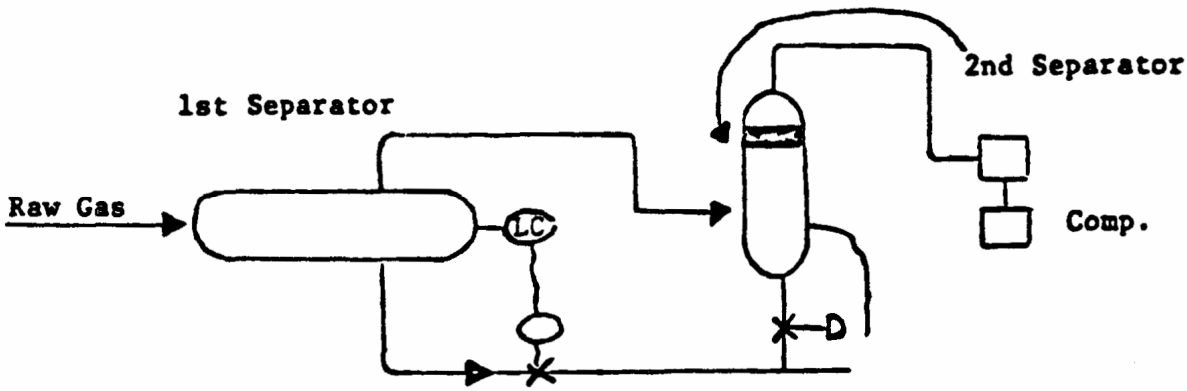
15.9 Other Types of Impingement Separators

Often baffles or packed beds are used instead of wire mesh. Equations are not generally available and manufacturer's data has to be used to size them. 'Brink et al' discusses some of these. Also, there are several other patent types of separators.

15.10 Problems

Example Problem 15.3

Below is the flow scheme for receiving wet gas from a field gathering system.



You are to determine the diameter and length of each separator.

- The gas flow is 100 MMSCFD.
- The gas pressure is 500 psia.
- The gas temperature is 70ºF.
- The gas has a molecular weight of 18.5.
- The liquid density is 38.9 lbs/cu. ft. - assume the same conditions in the second separator.

The system has to handle an increase of liquid of 100 bbls over 1 hour.

References

1. R.N. Watkins, "Sizing Separators and Accumulators, "Hydrocarbon Processing, Vol. 46, No. 11, November 1967, P. 253.
2. J.A. Brink Jr., W.F. Burggrabe and L.E. Green, "Mist Removal from Compressed Gases", Chemical Engineering Progress, Vol. 62, No. 4, April 1966, P. 60.
3. G..A. Gregory and K. Aziz, Journal of Canadian Petroleum Technology, July-September 1975, P. 29.
4. Wu, F.M. "Drum Separation Design, A new approach Chemical Engineering, Chemical Engineering, April 2, 1984.
5. Chapter 7, GPSA Data Book, 10th Edition 1987.
6. A. Baker, L. Nielson and A. Gabb, "Pressure Loss Liquid Holdup Calculations Developed", Oil and Gas Journal, March 14, 1988, P. 55; March 21, 1988, P. 78; March 28, 1988, P. 44.

16.0 Sour Gas Treating

- A review of the features, the design criteria, the operating problems and other pertinent data on the various methods of treating sour gas, ie. MEA, DEA, MDEA, HOT K_2C0_3, Sulphinol, Selexol. Propylene Carbonate, etc.

16.1 What is Sour Gas?

As far as a pipeline company supplying industrial or domestic consumers is concerned, any gas which contains greater than 1 grain of H_2S per 100 CF or 16 ppm (0.25 grains/100 CF or 4 ppm of gas for Pacific Coast to the United States) is sour. When the gas is used as a fuel for boilers, fired heater or for compressor engines, then gas containing up to 10 grains of H_2S/100 CF of gas is generally considered acceptable. Also, sometimes in remote locations compressors are driven by gas containing up to 2 % H_2S. Another case is the use of sour gas at wellhead as a fuel. This depends on the proximity of habitation and vegetation but 3-4 % H_2S content gas can be successfully used. The Alberta Energy Resources Conservation Board requires stacks for the flaring on start up or in emergency of wells in which the gas contains more than 1 % H_2S by volume.

Note: The CSA Standard Z184 Code for Pipelines defines sour gas as gas containing more than 1 grain of H_2S/100 CF (16 ppm).

Also, this section can be applied to the treatment of gas containing just CO_2. Although high contents of CO_2 are not nearly so undesirable, generally it is the practice to reduce the CO_2 content down to 2 % or whatever results in treating for 1125 removal as long as it is lower than 2 %.

16.2 Review of Processes

There are many processes for removing acid gases. The choice of a particular process is affected greatly by the specification required for the residue gas, the quantity of gas processed and the composition of the raw gas being treated. The various processes will be outlined and then described in more detail. It should be noted that in most processes, CO_2 is also removed but not in all. In fact a great deal of the technology for the removal of H_2S was developed first for CO_2 removal in the preparation of synthesis gas for ammonia production. Gas sweetening processes can be divided into five categories:

1. Chemical solvent processes.
2. Physical solvent processes.
3. Direct conversion processes.
4. Dry bed processes.
5. Miscellaneous Processes.

Table 16.1 gives a survey of the commercial gas sweetening processes; the list, however, is not complete, as new processes are constantly being introduced.

Table 16.1: Commercial Gas Sweetening Processes

I	Chemical solvent processes	Solvent

	A. Alkanolamines	**(typical)**
	MEA	2.5 N-Monoethanolamine (=15% in water)
	UCAR	5 N-Monoethanolamine (=30%) with inhibitor
	DEA	2 N-Diethanolamine (=21%)
	SNEA-DEA	3 N-Diethanolamine (=32%)
	DIPA	2 N-Diisopropanolamine (=27%)
	MDEA	2 N-Methyldiethanolamine (=24%) or higher %
	DGA (Econamine)	6 N-Diglycolamine (=63%)
	Inhibited Amines	(both MEA and DEA (30-50%)
	Hindered Amines	varying amines
	B. Hot Potassium Carbonate	
	Catacarb	pot.carb.solution with catalyst
	Benfield	pot.carb. solution with catalyst
	Giammarco-Vetrocoke	pot.carb. solution + arsenic trioxide
	Alkacid-M	pot.salt of methyl amino propionic acid
	Alkacid-DIK	pot.salt of dimethyl amino acitic acid
II	**Physical solvent processes**	**Solvent**
	Selexo 1	Dimethyl ether of polyethylene glycol (DMPEG)
	Fluor solvent	Propylene carbonate
	Purisol	N-methyl-pyrrolidone (NMP)
	Rectisol	Methanol
	Sepasolv MPE	Oligoethylene glycol methyl isopropyl ethers
	Sulfinol	Mixture of sulfolane, DIPA and H_2O
	Aminsol	Mixture of MEA, DEA with methanol
III	**Direct conversion processes**	**Solvent**
	Stretford	Sodium carbonate, sodium vanadate anthraquinone disulfonic acid and traces chelated iron
	Takahax	Sodium carbonate and 1.4 naphthoquinone, 2-sulfonate sodium
	Thylox	Sodium or ammonium thioarsenate solution Ferrox Sodium carbonate and ferric hydroxide solution
	Manchester	Sodium carbonate and ferric hydroxide solution
	Perox	Aqueous ammonia solution with hydroquinone
	Townsend	Triethylene glycol with dissolved
	Fugapol	Polyethylene glycol with metal salt catalyst
	G.V. -sulphur	Alkali arsenates and arsenites solution
	Lo Cat	Chelated iron in solution
	Sulferox	Chelated iron in solution
IV	**Dry-bed processes**	**Packing**
	Iron sponge	Iron oxide
	Zinc oxide	Zinc oxide
	Mol sieves	Molecular sieve
	Haines	Molecular sieve

V	Miscellaneous process
	H_2S Scavenging process
	These are chemical processes that are used to remove H_2S on a non-regenerable basis. The chemicals used include formaldehyde, iron oxide, zinc oxide, sodium hydroxide, sodium nitrite. For each use it is necessary to go to the literature.
	Membrane processes

Some of the more important processes are summarized in the next few paragraphs and then in the next section; some of the processes are described in some detail.

Also, a comparison of the fundamental difference between the various processes is given in Table 16.2

Table 16.2: Characteristics of chemical and physical solvents, direct conversion and dry bed processes

	I Chemical solvent	II Physical solvent	III Direct conversion	IV Dry bed
H_2S removal principle	chemical absorption	physical absorption	chemical conversion	a. chemical adsorption b. physical adsorption
H_2S loading	 loading limited by stoichiometry	 loading proportional to part. press. H_2S	 loading limited by stoichiometry	 loading limited by a. stoichiometry b. surface area
Quantity of H_2S to be removed	large	very large	small	very small
Purity required	moderate/high	high	moderate/high	a. very high b. high
Desorption energy	high	low	moderate	a. not regenerable b. moderate
Typical application	general purpose continuous	bulk removal continuous	continuous	a. guard bed (batch) b. cyclic operation

16.2.1 Dry Box Treating

H_2S only is removed by passing the gas through a bed of iron oxide and wood shavings. The reaction is as follows:

Reaction 16-1

$$3\,H_2S + Fe_2O_3 \rightarrow Fe_2S_3 + H_2O$$

Air is used to regenerate the bed.

16.2.2 Caustic Treating

H_2S is removed by contacting the gas with a caustic soda solution. The following reaction takes place:

Reaction 16-2

$$2\,NaOH + H_2S \rightarrow Na_2S + 2\,H_2O$$

The solution is dumped when it is spent.

16.2.3 Amine Treating

In this process an amine is used to react with the H_2S. The most common amines used are monoethanol amine ($NH_2\,C_2\,H_4\,OH$ or MEA) and diethanol amine ($NH(C_2\,H_4\,OH)_2$ or DEA). The reaction that takes place can be described by the following equation:

Reaction 16-3

$$RNH_2 + H_2S \rightarrow R\,NH_3HS$$

16.2.4 Hot Potassium Carbonate

In this process the acid gases are absorbed by a hot solution of K_2C0_3. The following equation represents the reaction:

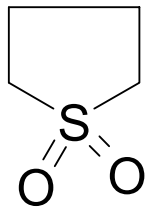
Reaction 16-4

$$K_2CO_3 + CO_2 + H_2O \rightarrow 2\,KHCO_3$$

There are several variations of this process. These include the "Catadsorb" in which an undisclosed catalyst is added to the carbonate solution and the Giammaro Vetricoke which uses arsenic compounds to activate the carbonate solution.

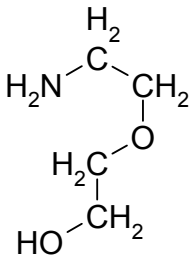
16.2.5 Sulphinol

In this process a mixture of a chemical called sulfolane and di-isopropyl amine (DIPA) along with water is used as the absorbing agent. The structure formula for sulfolane is:



16.2.6 Diglycol Amine

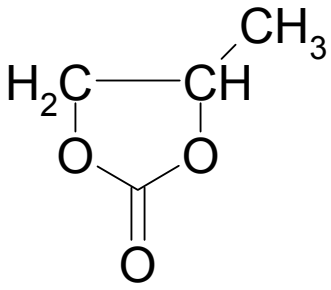
In this process diglycol amine (About 65-70 wt % in water) is used as the absorbing agent. Diglycol amine's formula is:



16.2.7 Other Chemicals

Other chemicals have also been proposed and used as treating agents commercially. These include:

a) Propylene Carbonate



b) Selexol Process - dimethyl ether of polyethylene glycol is the solvent.

c) The purisol process - N methylpyrrolidone is the solvent.

16.2.8 Miscellaneous Processes

These include some special methods such as:

a) Water Wash - In this method, high pressure water is used for physically absorbing the acid gases.

b) Low Temperature Condensation — Liquid H_2S is condensed out by means of low temperatures.

c) Townsend Process — In this process the gas is dried with glycol and then the H_2S is reacted with SO_2 to form elemental sulphur.

d) Ammonia Solution — This can be used for small concentrations.

e) Molecular Sieves - These desiccants have been proposed and used for treating low concentration of sour gases.

f) The Stretford Process — In this process in which elemental sulphur is produced directly, it uses Na_2CO_3, the sodium salt of anthraquinone disulphonic acid and sodium vanadate.

g) The Rectisol Process — The process uses cold methanol as an absorbent. The process is primarily used in synthesis gas application or the clean up in LNG plants.

Note: Good reviews of the various processes that can be used are those of Goar, Maddox, Tennyson Schaff, Lagas, Mohr and Ranke and King et al.

Table 16.3 indicates some physical properties of gas treatment chemicals.

Table 16.3: Physical Properties of Gas Treating Chemicals

	MONOETHANOL-AMINE	DIETHANOLAMINE	TRIETHANOLAMINE	DIGLYCOLAMINE®	DIISOPROPANOL-AMINE	SELEXOL®	PROPYLENE CARBONATE
Formula	$HOC_2H_4NH_2$	$(HOC_2H_4)_2NH$	$(HOC_2H_4)_3N$	$H(OC_2H_4)_2NH_2$	$(HOC_3H_6)_2NH$	Dimethylether of polyethylene glycol	$C_3H_4CO_3$
Molecular Wt	61.08	105.14	148.19	105.14	133.19	280	102.09
Boiling point @ mm Hg, °F	338.9	516.2	680 (Decomposes)	430	479.7	518	467
Freezing point, °F	50.9	82.4	72.3	9.5	107.6	−20	−56.6
Critical constants							
Pressure, psia	868	474.7	355	547.11	546.8	–	–
Temperature, °F	662	827.8	957.7	756.6	750.6	–	–
Density @ 20°C, gm/cc	1.018	1.095	1.124	1.058 @ 60°F	0.999 @ 30°C	1.031 @ 77°F	1.2057
Weight, lb/gal	8.48 @ 60°F	9.09 @ 60°F	9.37 @ 68°F	8.82 @ 60°F		8.60 @ 77°F	
Specific gravity 20°C/20°C	1.0179	1.0919 (30/20°C)	1.1258	1.0572	0.999 @ 45°C/20°C	–	1.203
Specific heat @ 60°F BTU/lb/°F	0.608 @ 68°F	0.600	0.70	0.571	0.69 @ 30°C	0.49 @ 41°F	0.335
Thermal conductivity, Btu/[(hr·sq ft·°F)/ft] @ 68°F	0.148	0.127	–	0.121	–	0.11 @ 77°F	0.12 @ (?)°F
Latent heat of vaporization, BTU/lb	355 @ 760 mm Hg	268 @ 73 mm Hg	230 @ 760 mm Hg	219.14 @ 760 mm Hg	185 @ (?)mm Hg	–	208 @ 760 mm Hg
Heat of reaction, BTU/lb of Acid Gas							
H_2S			−400	−674	–	−190 @ 77°F	–
CO_2			−630	−850	–	−160 @ 77°F	
Viscosity, cp	24.1 @ 68°F	350 @ 68°F (at 90% wt. solution)	1013 @ 68°F (at 95% wt. solution)	40 @ 60°F	870 @ 86°F 198 @ 113°F 86 @ 129°F	5.8 @ 77°F	1.67 @ 100°F 19.4 cs @ −40°F 1.79 cs @ 100°F 0.827 cs @ 210°F
Refractive index, N_D 68°F	1.4539	1.4776	1.4852	1.4598	1.4542 @ 113°F	–	1.4209
Flash point, COC, °F	200	280	365	260	255	304	270

	METHYLDIETHANOL AMINE	SULFOLANE®	METHANOL
Formula	$(HOC_2H_4)_2NCH_3$	$C_4H_8SO_2$	CH_3OH
Molecular Wt	119.16	120.17	32.04
Boiling point @ 760 mm Hg, °F	477.0°F	545	148.1
Freezing point, °F	−9.3°F	81.7	−143.8
Critical constants			
Pressure, psia		767.3	1153.9
Temperature, °F		1013.8	464
Density @ 20°C/20°C			
Weight, lb/gal	8.68	10.823 @ 30°C/30°C	0.7917
Specific gravity 20°C/20°C	1.0418	1.268	
Specific heat @ 60°F BTU/lb/°F	0.535	0.35 @ 30°C	0.59 @ 5°C−10°C
Thermal conductivity, Btu/[(hr·sq ft·°F)/ft] @ 68°F	0.159	0.114 @ 100°F	0.124
Latent heat of vaporization, BTU/lb	204	225.7 @ 212°F	474 @ 760mmHg
Viscosity, cp	1.3 cs @ 50°F 0.66cs @ 100°F 0.28 cs @ 212°F	10.3 @ 86°F 6.1 @ 122°F 2.5 @ 212°F 1.4 @ 302 0.97 @ 392°F	0.6 @ 68°F
Refractive index, N_D 68°F	1.469	1.481 @ 86°F	1.3286
Flash point, COC, °F	265°F	350	58

16.3 Process Descriptions

16.3.1 Iron Sponge

The gas to be treated is passed through a bed of sponge iron oxide and wood shavings. The chemical reaction that occurs has already been given. The bed is regenerated by either taking it off line and passing air through it or by continuously injecting a small stream of air in with the feed gas. The oxygen in the air reacts with the Sulphide and forms elemental sulphur and returns the iron to the oxide. Periodic regeneration is considered much better than continuous operation. This process is normally used at low pressure but has been used successfully at high pressures (See article by G.L. Duckworth and J.H. Geddes) and for gases containing less than 50 grains of H_2S per 100 cubic feet.

A few points to note about the process are:

The bed should be kept wet at all times and the pH should be kept above 8.0.

Reaction occurs best at temperatures between 80ºF and 100ºF.

When regenerating temperatures should be kept below 114ºF and the bed should be wet to prevent spontaneous combustion.

16.3.2 Caustic Treating

This process is only used as a clean-up as the products formed are not regenerable. The process consists of passing the gas to be treated through a 15% to 20% solution of NaOH in a vessel. The contents of the vessel are dumped when the caustic is consumed. This is determined by checking the effluent gas.

16.3.3 Amine Treating

There are two amines that are used extensively for sour gas treating; Monoethanol Amine (MEA formula $NH_2C_2H_4OH$) and Diethanol Amine (DEA, Formula $NH(C_2H_4OH)_2$). However other amines such as MDEA formula $CH_3N(C_2H_4OH)_2$ and hindered amines are being considered. We will discuss the MEA process first and then note the difference when DEA is used then discuss the other amine processes.

The properties of the MEA and DEA are given in Table 2.

16.3.4 The MEA Process

16.3.4.1 Process Description

Figure 16.1 shows the process flow sheet for a typical MEA sweetening facility. The sour gas at high pressures after passing through an inlet scrubber is led to an absorbing tower in which it is contacting with a 15%-20% solution of MEA in water. The acid gas components are absorbed into the amine solution and the gas passes out of the tower sweet. The rich amine solution is reduced in pressure to an intermediate flash tank where most of the absorbed hydrocarbons are flashed off and sent to fuel. The rich amine is then heated and flows to a still where the acid gases are stripped out by means of heat; generally steam is the heating medium. The acid gases are then sent to a sulphur plant or, if not much H_2S is present, just incinerated and vented up a stack. The lean amine from the still is cooled and recirculated back to the absorber.

There are two special slip streams of amine that are specially processed. The first is one that is often taken off the still. This is for reclaiming combined amine degradation products. It is a semi batch process. The other is a slipstream of lean amine for removing solids. These are both shown on the flow sheet.

16.3.4.2 Design Criteria

One of the principal design criteria is the circulation rate of amine. This for MEA for years has been set at a 3 moles of amine per mole of acid gas. This number has been based on the fact that at lower rates bad corrosion has been experienced. Also, it was based on interpreting the reaction equation as follows:

Reaction 16-5

$$2RNH_2 + H_2S \rightarrow (RNH_3)_2S$$ rather as is now assumed.

$$RNH_2 + H_2S \rightarrow RNH_3 - HS$$

The CO_2 reaction is generally considered to follow mostly:

Reaction 16-6

$$2RNH_2 + CO_2 \rightarrow RNHCOO^- + RNH_3^+ \quad carbonate$$

Also at greater concentrations of CO_2 than 0.5 mole CO_2/mole of amine the bicarbonate ion often forms:

Reaction 16-7

$$CO_2 + RNH_2 + H_2O \rightarrow HCO_3^- + RNH_3^+ \quad carbonate$$

rather than the equation given previously. Some plants have been designed to 2.5 moles of amine/mole of acid gas but many commercial plants are still operating at the 3 to 1 ratio. However, there does not seem to be any good theoretical reason not to go to the lower circulation rates that have proven successful with diethanol amines (DEA). One plant has been operating successfully a 1.6 mole amine to acid gas ratio.

The solution strength is normally set at 20% by weight of the treating solution could be used as is done in DEA Treating. See the discussion later on inhibited amines.

The number of trays to be used in the primary contactor is another important design variable. The common industrial practice is to use from 20 to 25 trays in the absorber. If a calculation of the theoretical number of trays is done from published equilibrium data, a value of 3 or 4 will be found. Thus if an efficiency of 35% for a tray is taken as absorption is taking place, then 12 actual trays would be required. Fitzgerald and Richardson shows that the first 10 trays pick up nearly all the H_2S and the last 15 trays are of not too much value. Figure 16.3 shows this data. It is suggested that for making 1 grain specification gas an amine contactor with 15 trays should be adequate.

Another very important design variable is the amount of steam required for stripping. The requirements for this have also been worked out by Fitzgerald and Richardson. They have developed a correlation based on the ratio of H_2S to CO_2 in the feed gas to the sweetening unit. This data is given in Figure 16.4. You will note that the required steam rate per U.S. gallon of solution circulated increase from about 0.80 lb to over 1.2 lbs as the H_2S to CO_2 ratio increases.

The number of trays in the stripper in the amine system is also of prime importance. A theoretical calculation of the number of trays required for stripping is given in Kohl & Riesenfeld, page 59 which shows that only 8 trays are required for good CO_2 removal.

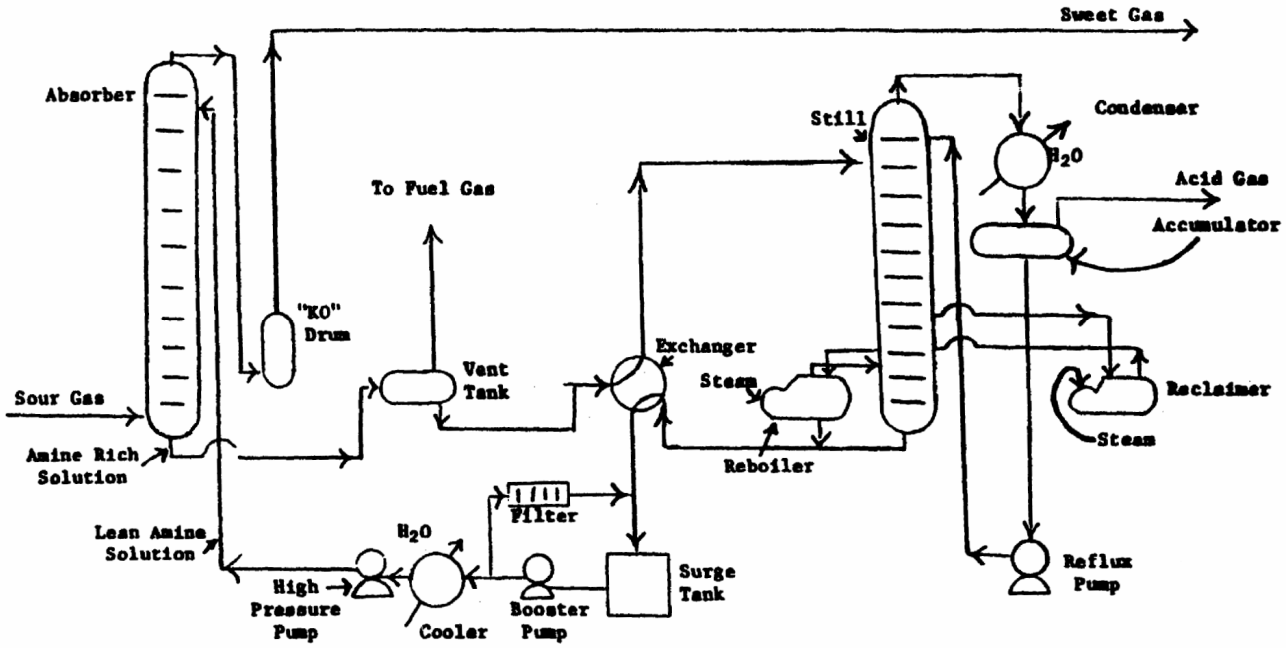


Figure 16.1: Typical MEA Sweetening Process

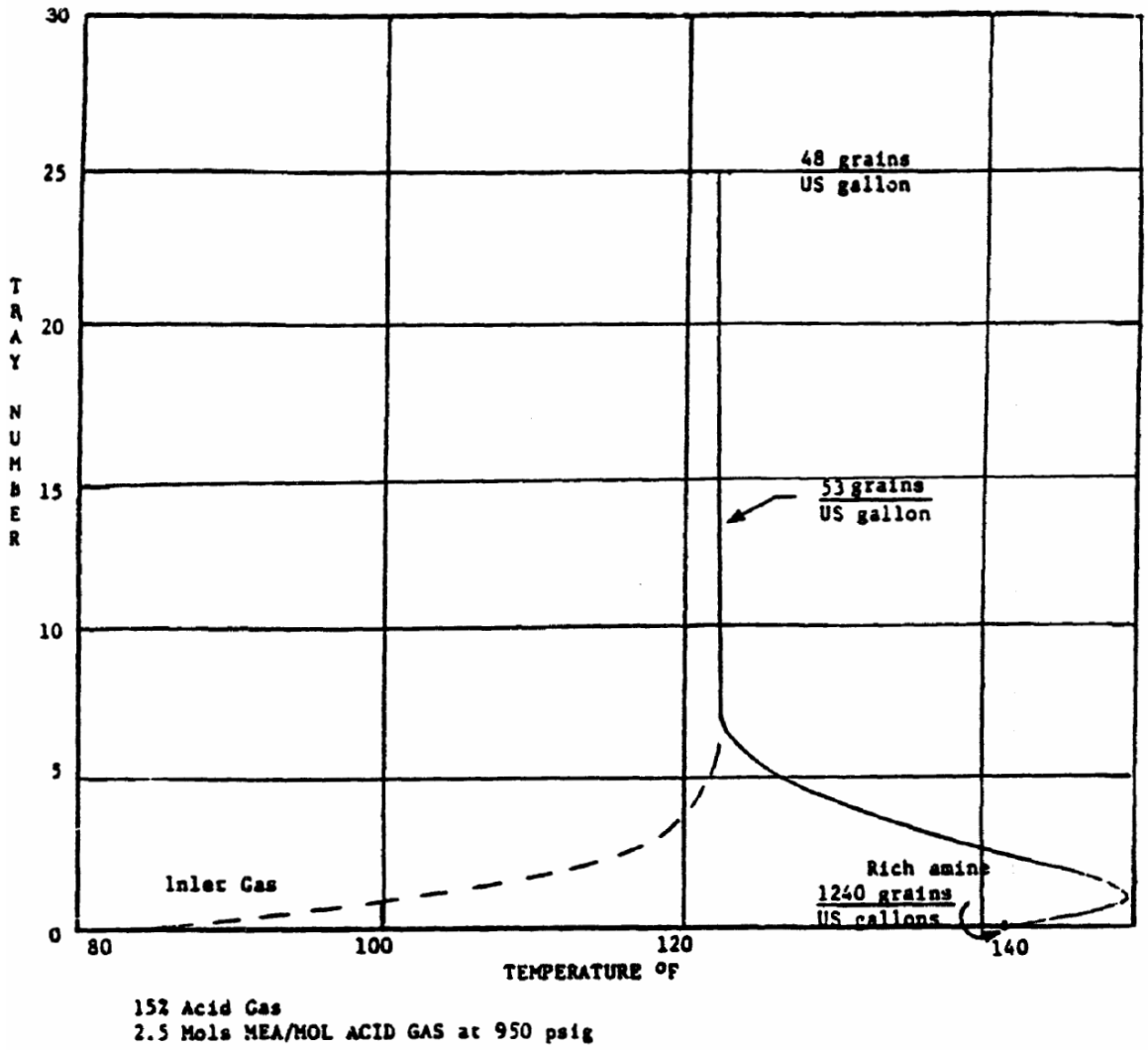


Figure 16.2: Typical MEA Contactor Temperature Profile

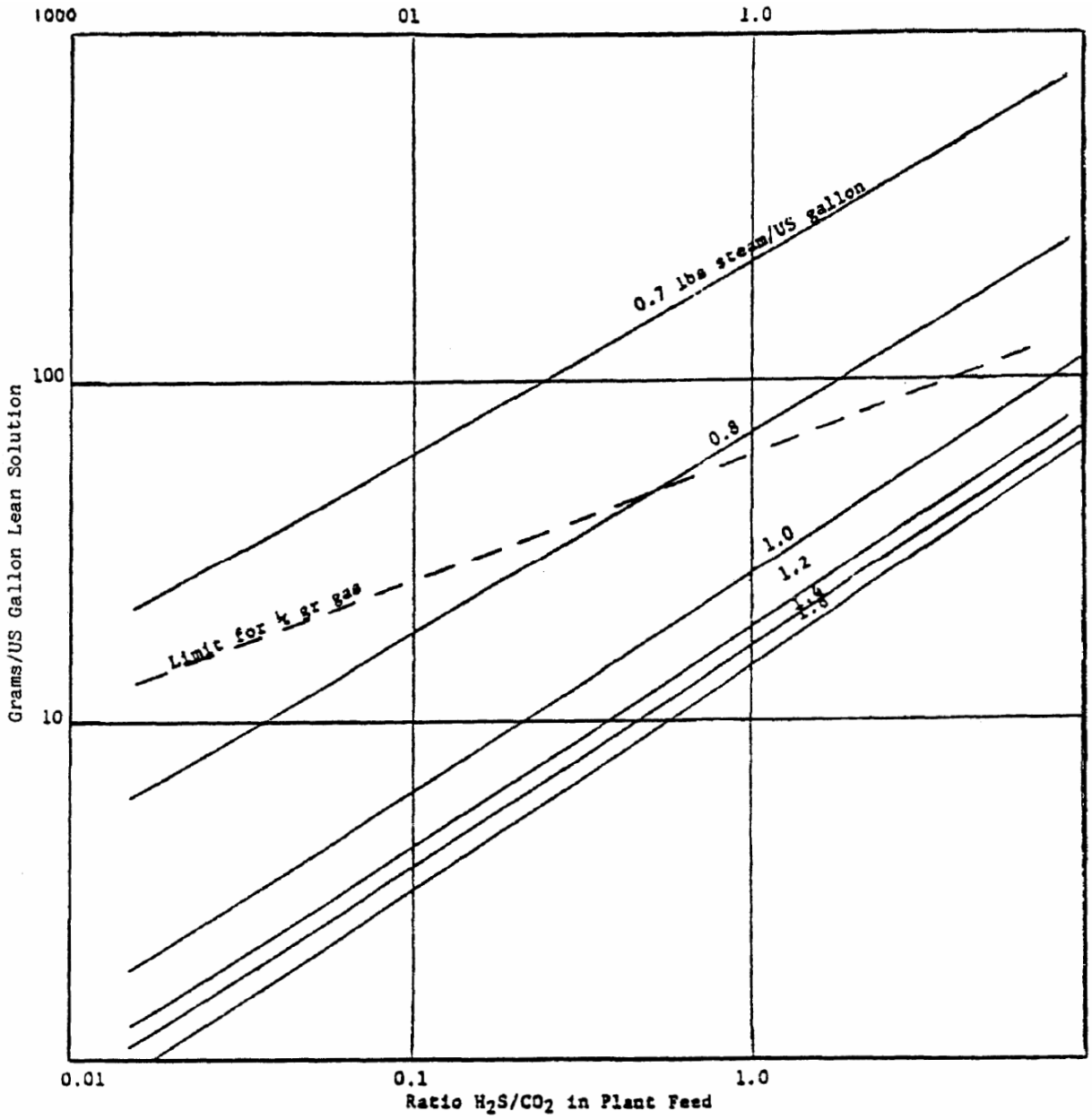


Figure 16.3: Steam Requirements for MEA Solutions per Fitzgerald and Richardson

Common industrial practice is for the column to have 20 trays. Other design criteria are that the reclaimer should have a feed rate of about 2% of the circulation of amine and a filter that handles between 5 and 10% of solution should be provided.

16.3.4.3 Operations

The plants that use MEA solutions generally have a few main problems. These are:

1. Foaming of the Solution - This can be minimized by keeping the solution clean primarily by good filtering and by keeping liquid hydrocarbons out of the solution. It can be controlled by the injection of antifoams such as Ocenol, Antifoam B and others. Provision for both continuous injection by a pump and slug injection by means of shotpots should be provided.
2. Corrosion - Corrosion occurs primarily in the following places:

a) Amine still reboilers.

b) Bottom section of amine stills.

c) Just after control valves in which rich solution has been allowed to flash.

d) Sometimes in the lean to rich exchangers.

Corrosion can be controlled by (a) proper design - keeping velocity of fluid down, (b) the addition of corrosion inhibitors, (c) use of coatings such as aluminum, gunnite, etc., (d) by keeping the reclaimer in operation to keep the degradable products from building up.

3. Loss of Solution - Another operating concern is the loss of amine which is a very significant operating cost. This can be minimized by:

- Collecting all amine drains in a sump and reusing in the system.
- Installing mist extractors on the outlet of the amine contactors.
- Providing water wash sections after the amine contactors.
- Keeping the reclaimer in operation to prevent the buildup of nonregenerable compounds.
- Not having COS or CS_2 in inlet gas.

Usually in the normal operating plant the amine losses run from 3/4 to 1 lb per MMSCF of gas treated per % acid gas present in the feed gas. If compounds such as COS and CS_2 are present they react with the MEA to form nonregenerable compounds and increase the losses considerably.

16.3.4.4 Special Notes

There is one special variation of the process that should be noted. That is the split cycle on which a semi lean (partially regenerated) solution is used for initial pickup and a lean (fully regenerated) solution is used for the final pickup. This does save on steam but requires more pieces of equipment and has been known to cause more than usual amounts of corrosion (See E. Berlie).

16.3.4.5 Summary of Process

Advantages

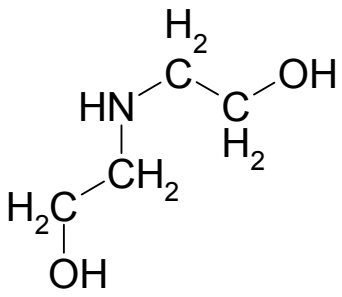
- Good pickup of acid gas.
- Well known and tried process (problems already known).
- Makes specification gas easily.

Disadvantages

- High capital cost compared to some newer processes because high circulation rate.
- High chemical losses with some impurities.
- Some corrosion problems.

16.3.5 The DEA Process

This is very similar to MEA treating except that the agent is diethanol amine instead of monoethanol amine. The chemical formula for DEA is:



This agent has been used for a long time in refineries primarily because it does not react with COS or other such impurities which are commonly present in refinery gases.

It has recently come into use for sweetening natural gas. This was started because of the high content of COS in some very sour gases, such as the ones at Windfall, etc.

In Lacq, France it is being used extensively. In fact a patent on the process is issued to the Societe Nationale des Petroles d'Aquitaine (SNPA) - Canadian patent 651,379. This patent describes the solution loading and steam rate required for this solution loading. An article by Wendt and Dailey describes SNPA Process applications. Younger and Kyllo reported the operation of a DEA plant above the patent range in loading.

An article by Smith and Younger reports on the performances of DEA plants in Canada. At present in Western Canada there are over 20 using DEA (Refer to Table 16.2). Eight of these use the SNPA process. To describe the process we will compare it with the MEA process:

16.3.5.1 Process Description

The process is very similar to the MEA one except that a reclaimer is not generally provided, as vacuum regeneration is required. This has been tried and not worked well. Thus the flow sheet is similar to Figure 16.1 with the reclaimer removed. Figure 16.2 shows the major equipment.

16.3.5.2 Design Criteria

The circulation rate is generally lower than for MEA. The reaction that is assumed to take place is:

Reaction 16-8

$$H_2S + R_2NH \rightarrow R_2NH_2^+ - HS^-$$

with the CO_2 reacting either to form the carbamate:

Reaction 16-9

$$CO_2 + R_2NH + H_2O \rightarrow HCO_3^- + R_2NH_2^+ \quad \textit{or the bicarbonate}$$

$$CO_2 + R_2NH + H_2O \rightarrow HCO_3^- + R_2NH_2^+$$

instead of the one given under MEA processing. Most DEA plants operate with a circulation rate based on 2.0 moles of amine, to 1 mole of acid gas or lower.

The SNPA patent covers the range of 0.9 moles to 1.3 moles of amine per mole of acid gas, with most plants being designed near the 1.3 ratio. This patent is no longer in effect. The article by Lee, Mather and Otto gives a way to calculate the minimum circulation rate.

The strength of the solution is also higher. Twenty to thirty weight percent of DEA is used.

The design of the amine contactor and amine still are the same as for MEA.

The design steam rate for DEA is the same as MEA and the data of Fitzgerald and Richardson can be used. In fact DEA seems to strip easier than MEA. However most people seem to use about 1.2 lbs/gallons of circulation. One special note is that DEA absorbs considerably more methane at high pressures so the rich solution vent tank should be designed accordingly.

Since there is no reclaimer, carbon is added along with the precoat or filter aid in the filter to absorb the degradation products. Most plants now use a carbon column with about 1% of the circulating amine passing through the column. The filter rate is generally 5 to 10% of the solution circulation as in the MEA process. Adequate filter capacity should be provided.

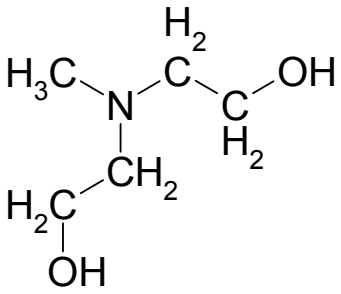
16.3.5.3 Operations

The operating problems are similar to those in MEA, i.e. foaming corrosion losses, etc. Foaming is generally less - water wash scrubbers are not required. Corrosion is similar to that experienced with MEA. Corrosion areas have been in the reboiler and in the rich solution side of the lean to rich exchanger when the velocity in the exchanger is high. Liquid losses are about 0.5 lbs of DEA/MMSCF of gas treated per.% acid gas.

Recent work by Union Carbide in developing corrosion inhibitor for both MEA and DEA have shown that the concentration of the amine can be increased to almost double, ie. 30-35% MEA and 50-55% DEA. The corrosion inhibitor makes the system passive. The processes are just the same as before except that much higher concentration of amines can be used. The paper by Kosseim et al gives particular data on four plants that were retrofitted using inhibitors.

16.3.6 The MDEA Process

The formula for MDEA is as follows:



The reagent is particularly selective to the adsorption of H_2S as the reaction previous described with H_2S:

Reaction 16-10

$$H_2S + R_2NCH_3 \rightarrow R_2NH^+CH_3 + HS^-$$

goes very quickly as before and since there is no H available to react to form the carbamate reaction the only CO_2 reaction is the formation of the bicarbonate which is quite slow.

Therefore if the circulation rate and contact time is properly selected the MDEA can essentially absorb all the H_2S and only a part of the CO_2. This principle was put into practice at Dome North Caroline plant as described by Daviet et al.

The particular features of MDEA besides its selectivity is it reduced corrosivity solutions up to 50% by weight can be used. The amount of heat for regenerating the MDEA is somewhat less than required for MEA and DEA (about 85% of the steam in lbs/gallon required for these amines). Also the degradation products formed are reported to be quite a bit less.

16.3.7 Hindered Amines

Exxon has developed a whole series of processes for particular applications[30] using special types of amines depending on the requirement for selectivity either for selective H_2S removal, high CO_2 removal, etc. A hindered amine is a compound that has a large organic molecular connected directly to the N in amine. Each type of amine has different selective and absorption capacities. Reduced circulation rates and reduced steam regeneration rates are claimed. Commercial units have been tested satisfactorily. At present it is used at the Quirk Creek plant operated by Esso.

16.3.8 Diglycolamine

This process is used extensively in the United States and in one plant in Canada.

The absorbing chemical is 2— (2-Aminoethoxy) ethanol with the formula:

$$NH_2\ CH_2\ CH_2OCH_2\ CH_2OH$$

or Diglycol amine.

The process flow sheet is very similar to MEA with essentially all the same units being used. The reclaiming is however primarily carried out in a vacuum.

The solution used is generally 65 to 70 wt % DGA. The design acid gas pickup is about 5-7 SCF acid gas/US gallon of solution.

The steam rate required is about 1.25 lbs of steam per gallon of solution circulated.

COS does not form a non-regenerable compound and thus gases having high COS can be handled.

Note: This process is described by Holder[10].

16.3.9 General

A good article on overall corrosion in amine plants is by Sheilan & Smith[34]. This article states that most corrosion is caused by local conditions which cause an upset in the equilibrium between the acid gases and the amine solutions particularly in areas of turbulence where free acid gas is liberated at high temperatures.

The following table shows heat of desorption.

Table 16.4: Heat of desorption

	H_2S (kJ/kg)	CO_2 (kJ/kg)
MEA	1905	1920
DEA	1190	1510
DIPA	1140	2180
MDEA	1050	1420

Although the heat of desorption is a small part of the total steam consumption, it has a beneficial influence, especially with MDEA.

The influence of the stripping steam is more important. Weak bases, like DIPA and MDEA, have a higher H_2S and CO_2 vapour pressure. The driving force of H_2S and CO_2 at regeneration is therefore higher, which reduces the stripping steam requirements. For selective absorption, steam savings compared to MEA, DGA and DEA, are normally 30-40%.

16.3.10 Hot Potassium Carbonate Treating

This process was developed at the U.S. Bureau of Mines. The article by Benson, Field and Jimenson reports on the process as it was applied to the removing of CO_2 from synthesis gas for NH_3 production. It has been used in three plants in Western Canada - the West coast Plant at Ft. Nelson, B.C., the I.O.L. plant at Nottingham, Saskatchewan, and the Canadian Occidental plant at East Calgary. Two of these plants have been converted to other solvents and one is shut down. To work successfully an amine clean-up section is required when treating gases containing H_2S.

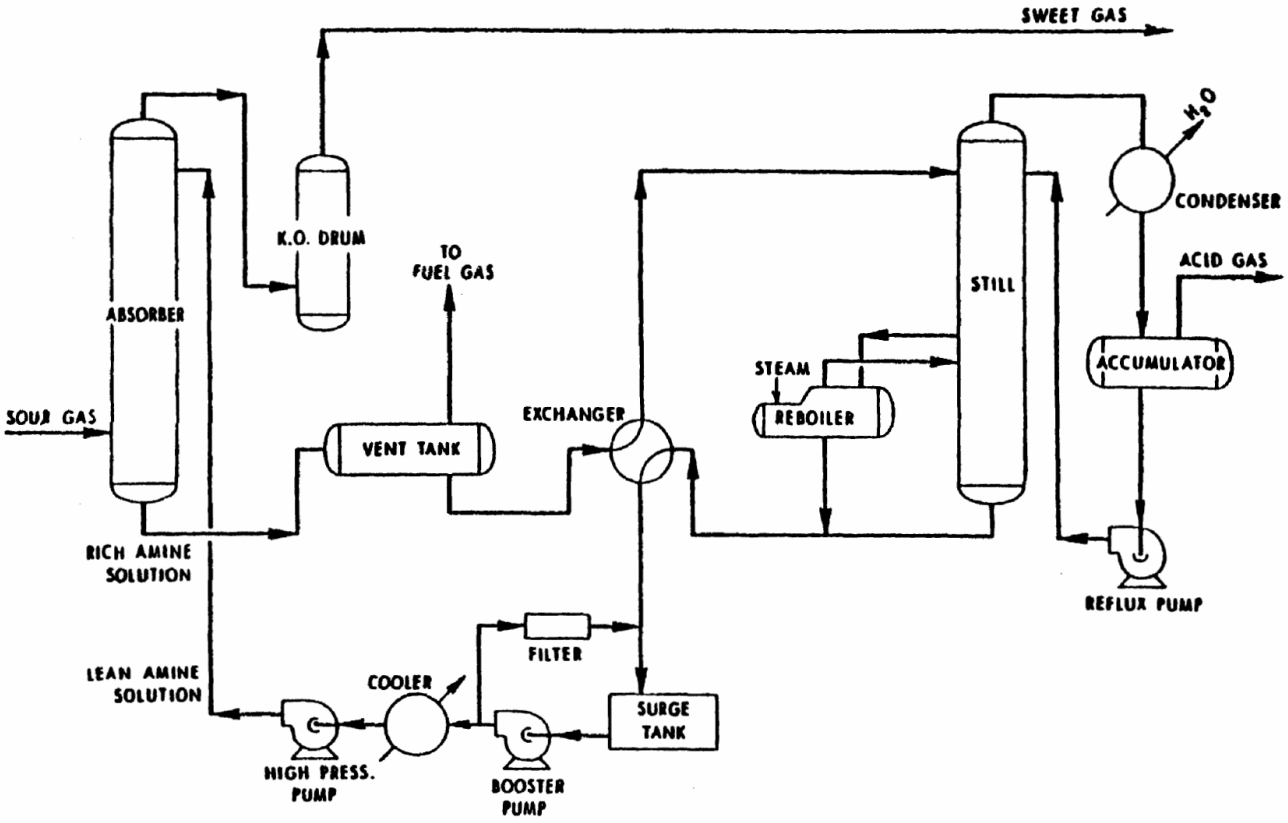


Figure 16.4: DEA Amine Sweetening Process

The hot potassium carbonate process is a good one for straight CO_2 removal. Most substitute gas manufacturing plants that have been built use this process for CO_2 removal.

The chemistry of the reactions that occur are as follows:

Absorption

Reaction 16-10

$$K_2CO_3 + CO_2 + H_2O \rightarrow 2\,KHCO_3$$

$$K_2CO_3 + H_2S \rightarrow KHCO_3 + KHS$$

Stripping

Reaction 16-11

$$2\,KHCO_3 \rightarrow K_2CO_3 + H_2O + CO_2$$

$$KHS + KHCO_3 \rightarrow H_2S \uparrow + K_2CO_3$$

The reaction of absorption goes better at high temperatures. Figure 16.5 shows this data.

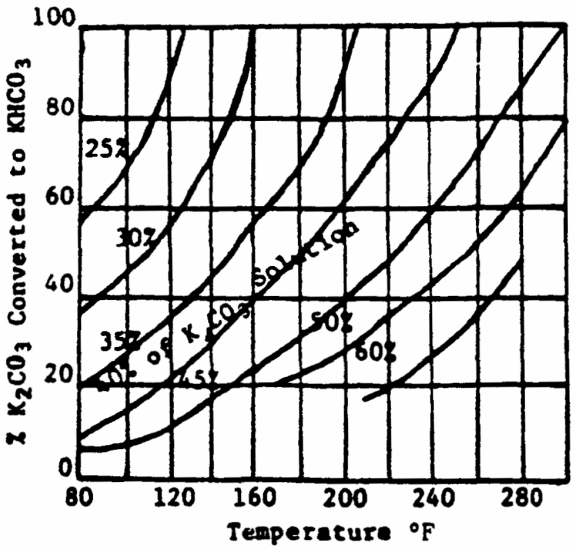


Figure 16.5: Reaction Rate K_2CO_3 Process

16.3.10.1 Process Description

This process is quite similar to that of MEA. Figure 16.6 shows a typical flow sheet for the process. You will note there is an inlet heat exchanger to warm up the incoming gas with effluent sweet gas from the contactor. Also there is no intermediate heat exchange between rich and lean solutions. A side stream filter is used to clean the solution of iron sulphide and various other impurities.

16.3.10.2 Design Consideration

a) Solution Strength - Equilibria data show that 40% and 50% weight K_2C0_3 can be used. These strengths, however, have caused the cutting out of pump cases, etc. Thirty percent is now considered the maximum strength to use for design.

b) Acid Gas Pickup - The commonly accepted design value is 3.0 cubic feet of acid gas per gallon of 30% solution or 2 moles of K_2C0_3 per mole of acid gas.

c) Steam Rate - A steam rate 0.8 lbs of steam per gallon of solution circulated has been found to be adequate.

d) Number of Trays in Contactor - This depends on how much of the acid gas is removed in the contactor. Experience has shown that it is difficult with H_2S present to make pipeline specification gas and there is usually installed an amine clean-up unit. In this case only 8-10 trays need be put in the contactor. Figure 16.7 is a typical graphical representation of how to calculate the number of trays. As is shown only 2 trays are required for the case considered by tray efficiencies are quite low (in the 15% to 25% range) and thus 8 to 10 trays or more should be used.

e) Number of Trays in Still - Experience has shown that about 20 trays should be used although theoretical calculations show only three or four are required, but again efficiency is low.

f) Filter Slip Stream Rate - This should be designed for 5% of the total K_2C0_3 solution circulation.

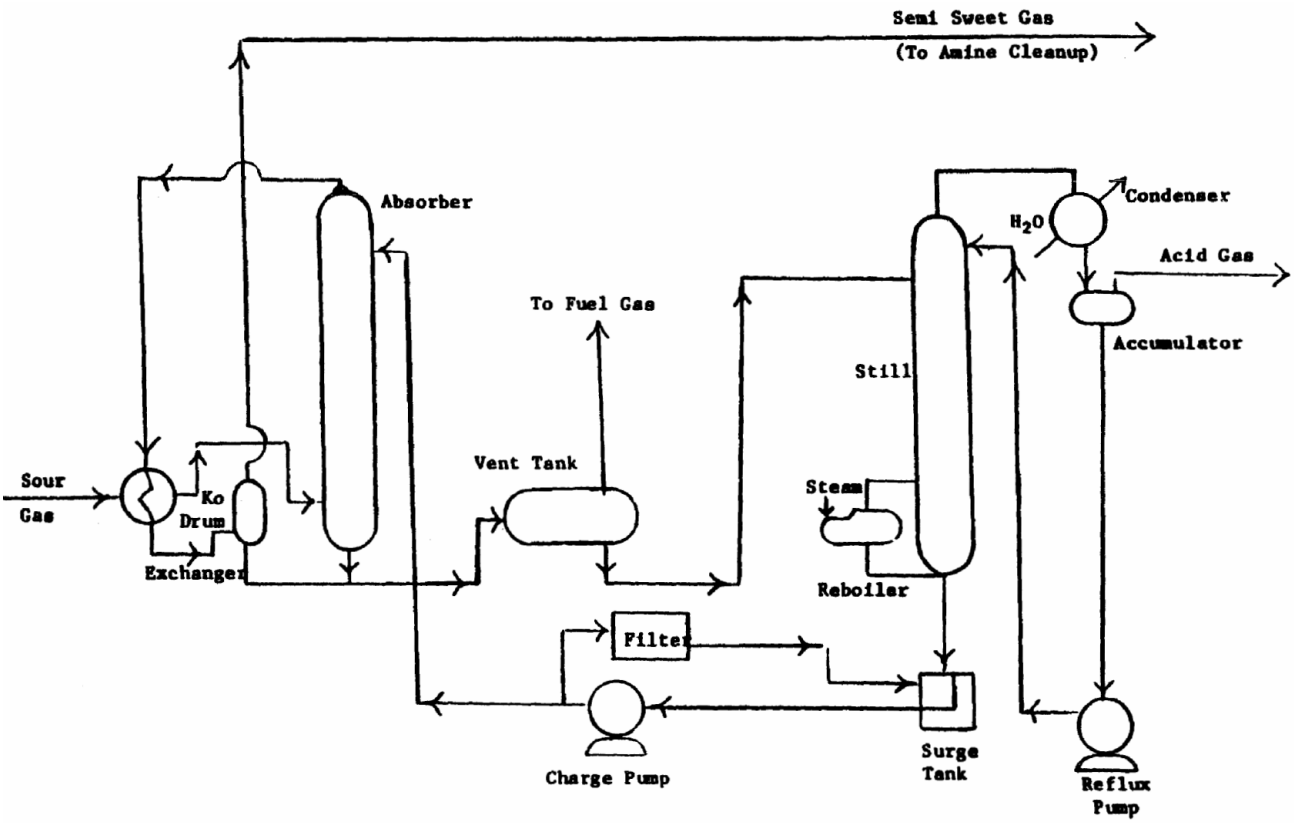


Figure 16.6: Typical Hot Carbonate Flow Sheet

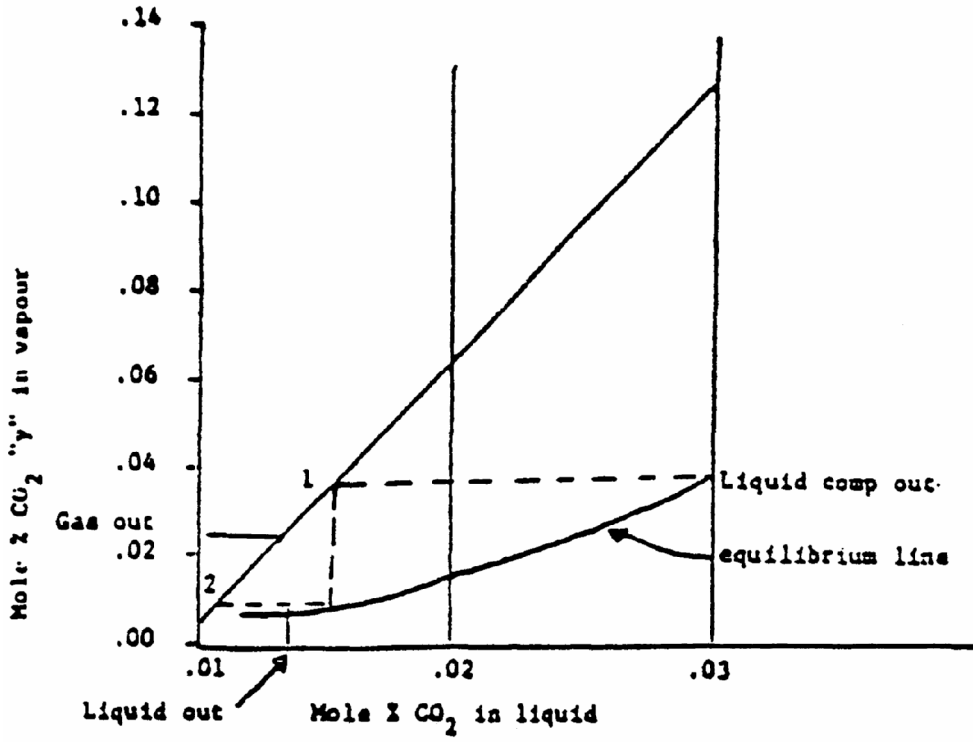


Figure 16.7: Tray Calculations for CO_2 Absorber

16.3.10.3 Operations

Some of the plants have experienced severe erosion problems. This has been generally controlled by keeping the solution strength below 30%.

Another problem is loading up of the solution near its equilibrium value with acid gas and the amine unit not being able to handle the sour gas that comes through. This situation can be improved by designing the carbonate section for full pickup of acid gas and the amine unit for a 10% gas pickup rather than 5% for the amine unit.

Chemical losses are generally quite small as the solution does not form degradable products, nor does it have an appreciable vapor pressure. Mechanical losses should run 0.5 to 0.6 lbs per MMSCF.

Corrosion can be quite severe but usually it is controlled by additives (potassium dichromate when no H_2S present and vanadium pentoxide when H_2S is present).

16.3.10.4 Special Notes

Gases that are treated should have a partial pressure of 20 psia or greater of CO_2 present. Also, ratio of $H_2S/C0_2$ of greater than 3/1 should not be treated. In general gases that have high CO_2 content should be considered.

There are a couple of patent versions of the Hot Pot process that use special chemicals in the solution. These are not as yet used in Canada but have been used in the U.S. and Europe.

16.3.11 Giammaro-Vetrecoke Process

This process was developed in Italy for removal of acid gases from coke oven and synthesis gases. A plant using this process is running in the U.S.

A discussion of the process is given by Riesenfled and Mullowney. The prime feature of the process is the use of arsenic compounds in the carbonate solution.

H_2S can be removed selectively by using arsenic compounds alone and then the CO_2 removed by carbonate with some arsenic and organic activators.

The advantages of the process for natural gas are as follows:

1. The increased solution activity for absorbing high amounts of acid gases and thus reduce solution circulation rate; also the height of the absorber.
2. Regeneration requirements are low; air can be used for regenerating arsenic compounds.
3. No corrosion problems.

The first U.S. plant is Trans Western Pipeline Co's near Ft. Stockton which treats a gas containing 28% CO_2 and 2 grains/100 cf H_2S. See Jennett.

Another patent process is the Cat Absorb Process developed by Eickmeyer which uses additives to a regular hot potassium carbonate system that increases the activity and reportedly negates the need for an amine clean-up. The additives are amine borates. It has been used primarily for synthesis gases but it is now being used for sour gas in Libya.

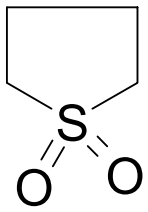
A good update on the Hot Potassium Carbonate process is given in the article by Bartoo.

16.3.12 The Sulphinol Process

A process that has found considerable acceptance is the "Sulphinol" process that was developed by Shell Oil (See article by C.L. Dun, E.L. Frietas, E.S. Hill, J.E. Sheeler). This process is now used in eleven plants in Western Canada. Refer to Table 16.2 for a listing. Goar reviews the process in some detail.

The solvent used is a mixture of:

- Sulfolane which has a structural formula of:



- Diisopropanol Amine (DIPA) formula $(HOC_3H_4)_2NH$.
- Water.

The exact quantity of each component varies but the sulfolane and the DIPA are usually in the range of 40 to 45% each and the water makes up the rest. The acid gas absorbing agent is primarily the amine. The sulfolane is a good absorber of all gases and improves the pickup of the acid gas. The DIPA constituent accomplishes chemical absorption as it combines with the sour components which is not sensitive to pressures. The sulfolane is a physical solvent that gives high absorption at high pressures. Thus an overall high absorption capacity is developed.

Table 16.5: Effect of Pressure on Sulfinol Circulation Rate

Item	Low H_2S content Gas "A"			Medium H_2S content Gas "B"			High H_2S content Gas "D"		
Contactor pressure, psia	100	500	1,000	100	500	1,000	100	500	1,000
Sulfinol solution circulation rate, gpm at 110ºF	2,000	1,540	1,483	3,300	1,995	1,748	6,410	3,230	2,366
Gas Content Mole %									
H_2S	0.65			2.0			51.5		
CO_2	8.73			1.4			3.5		

16.3.12.1 Process Description

The process is similar to the previous processes as described. A flow sheet showing the process is given as Figure 16.8. The primary difference is generally the need of additional flash drums on the rich sulfinol solution so as to release as much absorbed hydrocarbon gas as possible before the solution is regenerated.

16.3.12.2 Design Criteria

Specific design requirements should be obtained from the process licensor, the Shell Development Company, or an engineering contractor. Some general information is given below.

a) Solution Circulation Rate - Acid gas pickup of 6-17 Standard cubic foot/gallon of solution are used.

The contactor pressure has quite an effect on the required circulation as is shown in the next Table 16.4 (ie, lower pressure requires higher circulation). This is typical for all physical

solvents as indicated at the beginning of the chapter the pick up is a function of the partial pressure of acid gas.

b) Steam Requirement for Regeneration - A steam rate of 0.5 lbs per gallon of solution is generally found adequate.

c) Number of Trays in Contactors - General practice has been to use at least as many trays as used in an amine contactor, possibly slightly more (20-28)

d) Number of Trays in the Still - General practice has been to use the same as in amine service (15-24).

e) A reclaimer to handle about 1% of the solution is sometimes required. This is quite an involved system a tower, reboiler, vacuum pump, etc.

f) A filter to handle 5% of the solution is usually provided.

g) Often a two stage vent system is provided to remove absorbed hydrocarbon gases.

16.3.12.3 Operations

a) The solution has a good affinity for mercaptan and removes them.

b) Its foaming tendency is much less than amine.

c) No extensive corrosion has been reported; some in reboilers though.

d) It does remove COS and CS_2 and does not degrade nearly as much as ordinary amines.

e) It does absorb extensive amounts of hydrocarbons which, if they are not removed, will be sent to the sulphur plant and thus cause the formation of black sulphur. This is a particular problem with heavy hydrocarbons.

f) Losses of chemicals are quite expensive although the actual quantity used is low. They seem to run at 1.5 to 2.0 lbs/MMSCF of gas treated. Actually the DIPA and sulfolane should be looked at separately in this matter.

16.3.12.4 Special Notes

a) The gases that have been used for feed stock to sulphinol units are primarily very dry, with very little heavy hydrocarbons present. In particular aromatics should not be present in the gases.

b) The royalties charged for the process were quite high (Patent now expired).

c) The process should be considered for expanding existing plants because of high acid gas pickup.

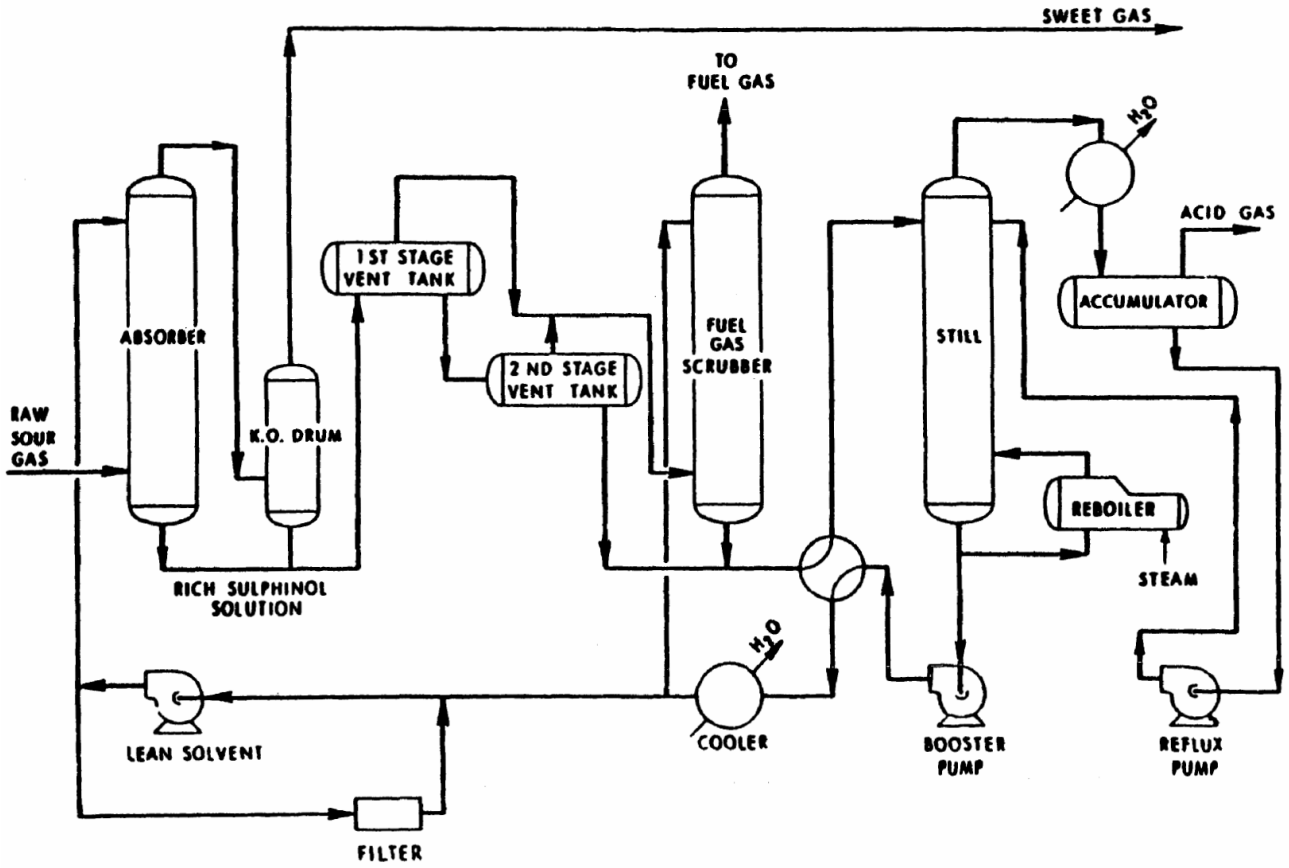
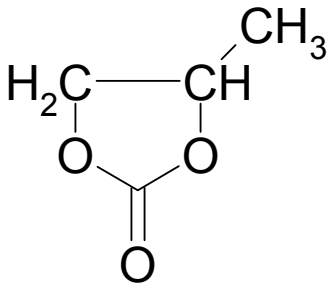


Figure 16.8: The Sulphinol Sweetening Process

16.4 Other Solvents

16.4.1 Propylene Carbonate (Fluor Solvent Process)

In this process the solvent is a chemical compound of the formula:



and is used to absorb the acid gases. The solvent has a high absorptivity of CO_2 and possibly H_2S although it is reported that it does not chemically react with them. It also reportedly does not absorb low mole wt. HC and H_2. The solvent was developed for synthesis gas units with CO_2 present but has been recently tried for sour gases (Low H_2S content). Buckingham reports on this process and how it is working.

The process is simpler than more conventional processes as regeneration is sometimes accomplished by lowering the pressure only, as shown in the flow sheet below.

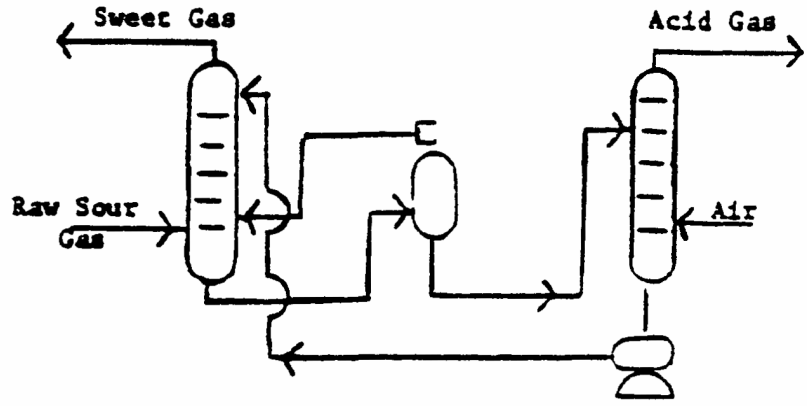


Figure 16.9: Propylene Carbonate Flow Sheet

16.4.1.1 Special Notes

a) Used on streams of high $C0_2$; partial pressure experience on gases containing high H_2S content not reported.

b) Reduces CO_2 to 1 to 2%. Would expect for pipeline gas with high H_2S might have to use amine clean-up.

c) For good regeneration probably need vacuum regeneration or steam.

16.4.2 Selexol and Other Similar Solvents

Another family of solvents that have developed for sweetening are those developed by Allied Chemical Co. and reported by G.S. Kutcher, G.A. Smith, D.A. Green and more recently by R.T. Clare and J.P. Valentine. The solution has a high capacity for acid gases (10.5 cf/gallon). In very sour gases dispropyl amine is added to the solvent which is a dimethyl ether of polyethylene glycol. Good removal of mercaptans in the gas stream is claimed. Also the solvent dehydrates the gas to pipeline gas specifications. This process is particularly good for gases high in CO_2 and low in H_2S.

16.4.3 Purisol Process

In this process n methyl 2—pyrrolidone is used as the solvent. The process is quite similar to the sulphinol process and is applicable to similar types of gases. It has been used in Germany for several years, K.J. Beavon and T.R. Rozkowki have described the process.

16.4.4 Miscellaneous Processes

There are other processes proposed. These include:

a) Water Wash - This system has been used extensively for the removal of CO_2 in synthesis gas and is now being proposed for H2S and CO_2 based on pilot plant data.

b) Low Temperature - Some pilot plant data has been reported for the removal of H_2S by freezing it out. U.K. Tin and F. Kurata discuss the theory of the process. Shell's plant at Waterton originally used this process.

c) Townsend Process - For H_2S removal primarily the process consists of reacting H_2S in sour gas with SO_2 by burning part of sulphur formed. Reaction takes place in tower in presence of Diethylene glycol which absorbs water formed. Also gas is dried further by contact with concentrated DEG. This process is good only for H_2S removal. Sulphur forms as a slurry and is filtered out.

d) Ammonia Solutions - The use of ammonia in the purification of gases from the cool gas process has been long practiced. It is not used in natural gas processing, Kohl and Rosenfeld discuss the process in chapter 4 of their book on Gas Purification (McGraw-Hill, 1960).

e) Molecular Sieves - The adsorption of H_2S on mole sieves has been tried and reported successful in certain cases and not in others. The principal is the selection of the proper size of sieve to absorb the H_2S. The absorbed H_2S is driven off with heat. The process may be satisfactory at low H_2S contents but at higher values it has not been satisfactory, ie. 2% gas at Buick Creek and Jedney, etc. It has also been used by Amoco on fuel gas as described by Kunkel and Chobotuh. In this case gas containing 0.57% mole H_2S is sweetened.

f) Membranes - Membranes that separate gas are being considered extensively for bulk CO_2 and H_2S removal. Generally if pipeline quality gas (ie. 4 ppm H_2S maximum) is required on clean up amine unit as required.

g) Stretford Process - The Stretford process uses an aqueous solution of sodium carbonate, which reacts with the H_2S to form NaHS, as below:

Reaction 16-12

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

The hydrosulfide is oxidized to sulfur by sodium vanadate also the solution.

Reaction 16-13

$$NaHS + NaHCO_3 + 2NaVO_3 \rightarrow S + Na_2V_2O_5 + Na_2CO_3 + H_2O$$

Subsequently, the vanadium is oxidized back to the pentavalent state by blowing with air, with anthraquinone disulfonic acid (ADA) or sodium anthraquinone disulfonate, acting as an oxidation catalyst:

Reaction 16-14

$$Na_2V_2O_5 + \frac{1}{2}O_2 \xrightarrow{ADA} 2NaVO_3$$

The spent absorbent solution flows from the absorber to the regenerator (oxidizer), where air is sparged into the tower. Tiny particles of sulfur are collected as a froth at the top of the oxidizer tower. The sulfur froth is skimmed off the solution and goes down a duct to a filter or centrifuge for removal of the filtrate from the sulfur cake. The clear (regenerated) absorbent solution and the filtrate from the filter are recycled back to the absorber.

The sulfur cake may be washed, dried, melted, and added to the product from the parent Claus unit. The solid sulfur produced from the Stretford process is usually melted in an autoclave. A flow sheet of the process is given as Figure 16.10.

This process is primarily used with low concentration of H_2S, ie. 1 mole % or lower (recent reference is Wilson and Newell).

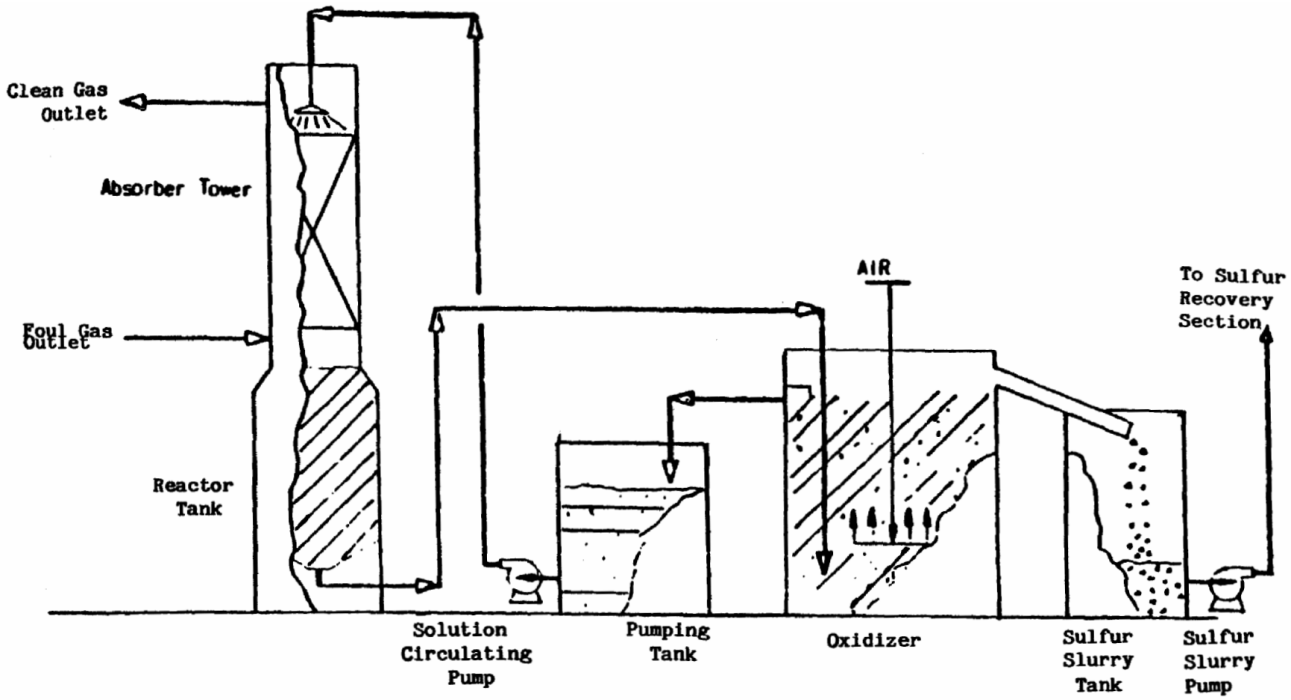


Figure 16.10: Stretford Process

h) Lo Cat Process - This process is a means of removal for small amounts of H_2S from large amounts of acid gas. It produces elemental sulphur initially as a slurry, then if a melter is installed, as a liquid. The sulphur formed meets commercial spec's except the colour is pale yellow or brownish yellow not a bright yellow.

The process consists of oxidizing the S ion to elemental $S^{=}$ by means of a metal ion usually Fe by the reaction indicated below:

Reaction 16-15

$$S^{=} + 2\,Fe^{+++} \rightarrow S^{\circ}_{(solid)} + 2\,Fe^{++}$$

Then reactivate the Fe^{++} to Fe^{+++} by the following reaction.

Reaction 16-16

$$\frac{1}{2}O_2 + H_2O + 2\,Fe^{++} \rightarrow 2OH^{=} + 2\,Fe^{+++}$$

Two vessels are required one for absorption of the H_2S and one for precipitation of the elemental sulphur and regeneration of the ion solution.

Approximately 30 commercial installations are operating four Alberta.

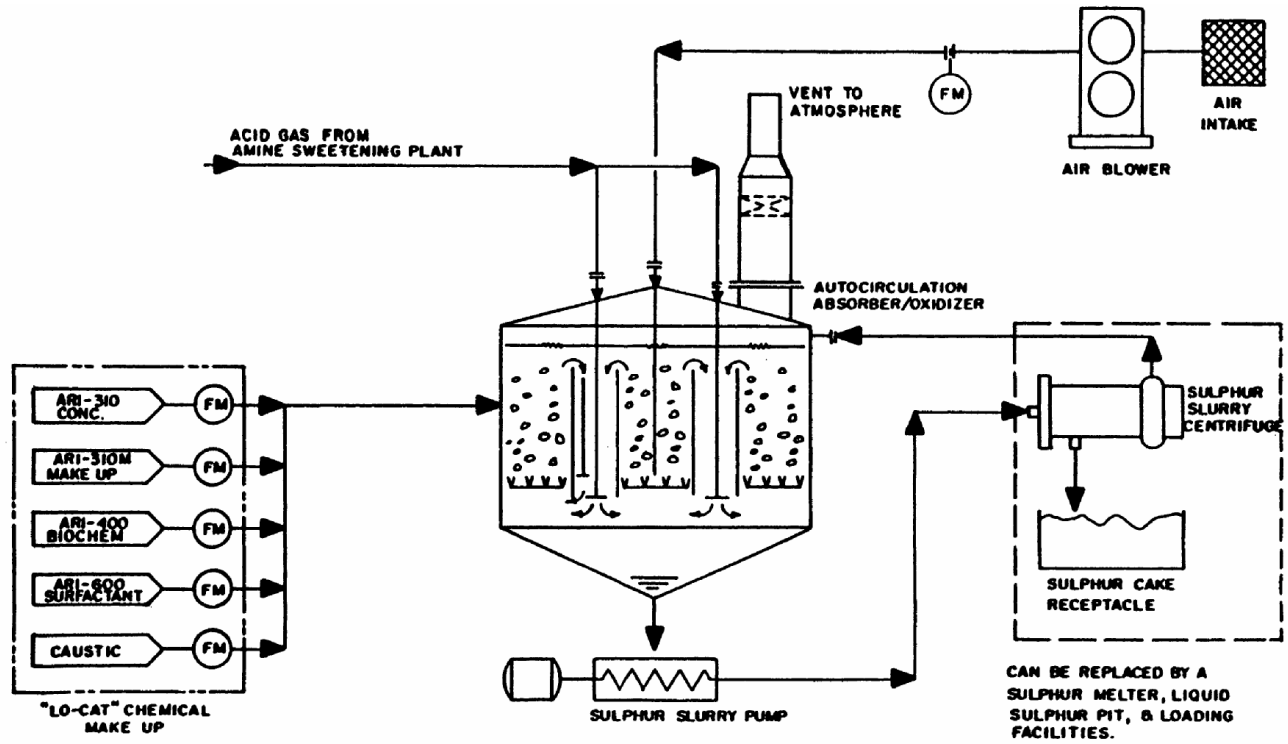


Figure 16.11: Lo-Cat Sulphur Recovery Process

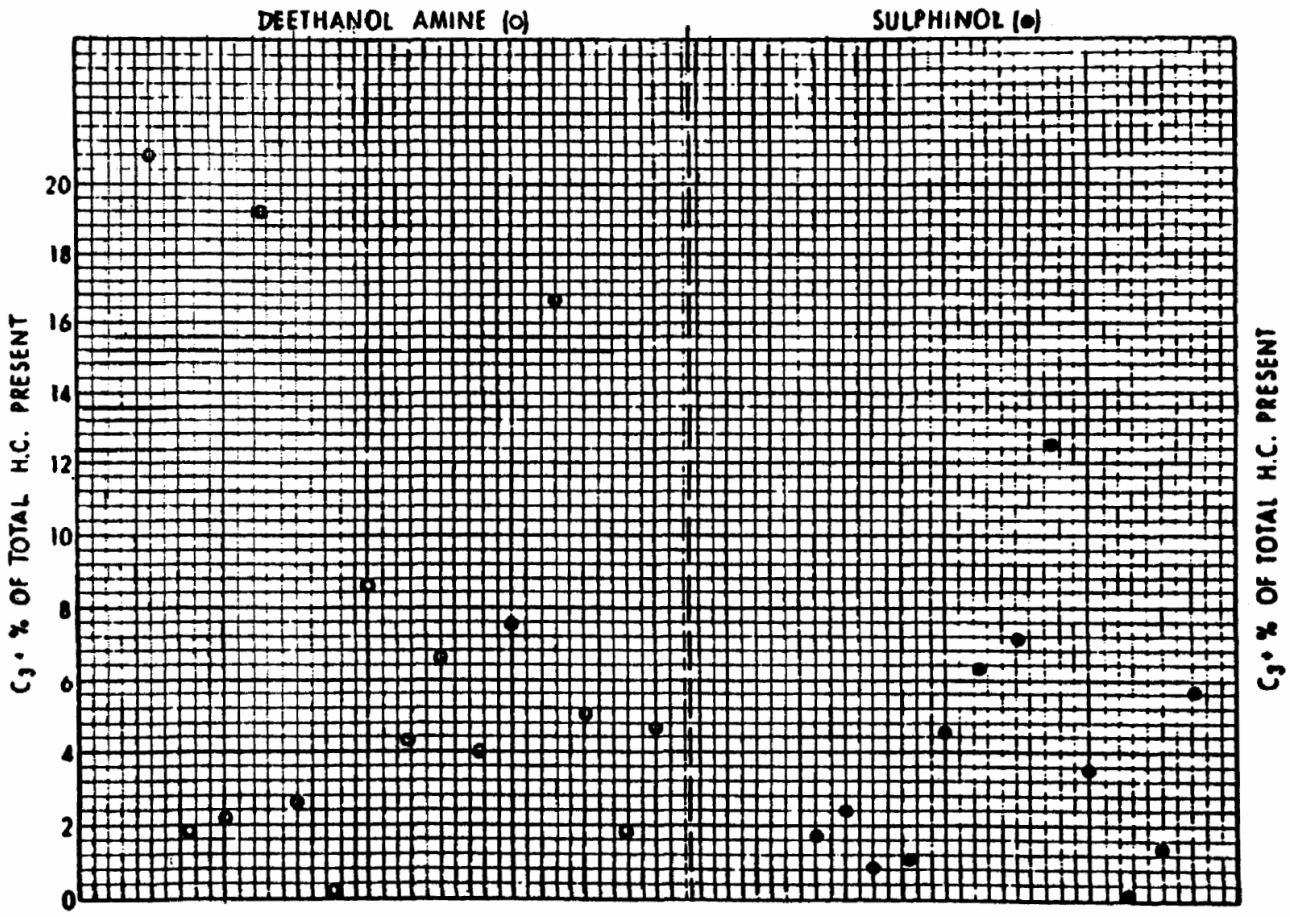


Figure 16.12: C_3^+ Content of Gases Treated by DEA and Sulphinol

16.5 Selection of a Sweetening Process

In recent years the two treating processes that have been picked primarily for use in Western Canada are either DEA or sulphinol, however, for special cases MDEA is being used. Younger, Lagas and others have discussed the choice of processes.

A rough way to decide on the process is as follows:

Sulphinol is used primarily on so-called dry gases, ie. when there is very little C_5^+ or even much C_3, C_4 present though it has been used on a gas from a crude stabilizer. Heavier components are removed first by refrigeration, in this case. It has also been used to remove mercaptans from the gas.

The primary disadvantage is that the heavy hydrocarbons are absorbed and are sent over to the sulphur plant (carbon adsorption has been tried but is reported to be not too successful).

DEA is used primarily when treating hydrocarbon rich gases, ie. high content of C_5^+.

A comparison of the heavier hydrocarbon content (C_3^+) of the gases that are treated by sulphinol and those that are treated by DEA are shown in Figure 16.11. As can be seen, the gases fed to sulphinol treaters are on the whole, much drier (i.e. have less C_3^+ content as compared to total hydrocarbon content) although the ranges do overlap. In the three cases in which sulphinol is the

treating agent and the gases have higher than 6% C_3^+ in the total hydrocarbons, refrigeration of the gas prior to treatment is practiced to get rid of the really heavy hydrocarbons.

Thus a general guide for the preliminary selection of the two processes might be as follows:

- less than 4% C_3^+ as a %HC sulphinol favoured.
- 4% to 6% C_3^+ generally a toss-up. In particular, the make-up of the C_5^+ portion of the heavy hydrocarbons should be investigated.
- above 6%, use DEA or other chemical agent.

Of course a detailed economic study should eventually be made generally following the method below.

16.5.1 Specific Selection Procedure

In selecting a gas sweetening process, the acid gases to be removed are divided into the following 4 groups:

a) CO_2 removal.

b) H_2S removal.

c) H_2S and CO_2 removal.

d) Selective H_2S removal.

1. In the first step a choice is made between a chemical, physical, direct conversion or dry bed process. The following points are taken into account:
   - Partial pressure of acid gas in the feed.
   - Product gas specification.
   - Impurities in the feed gas particularly the heavy hydrocarbon content and also such things as mercaptans and finally the expected further processing.
   - Quality of acid gas.
   - Quantity of sulphur to be removed.

Figure 16.13, Figure 16.14, Figure 16.15 and Figure 16.16 give guidelines on making an initial selection on the basis of partial pressure of the acid gas in the feed gas versus the product gas specification. The influence of impurities in the feed gas on the selection is outlined in 'Table 16.6: Maximum Practical Acid Gas Loading

Sweetening Agent or Process	Chemical Strength %	Solvent	Pick-up Moles Sweetening Agent/ Mole Acid Gas	Pick-up Acid Gas Scf/gal of Solution	Comments
MEA	20	H_2O	2 - 3		
MEA (with inhibitor)	30	H_2O	2 - 3		
DEA	25	H_2O	1.35 - 2		
SNPA - DEA	25	H_2O	0.9 - 1.3		
DEA (with inhibitor)	55	H_2O	1.35 - 2		
MDEA	50	H_2O	1.2 - 1.5		
DGA	50	H_2O	2 - 2.5		
Sulfinol	**	H_2O	----	6 - 17	** Sulfolane & DIPA

					90%.
Selexol	----	----	----	up to 10.5	See notes for chemical.
Hot Potassium Carbonate	30	H_2O	2 - 3		Difficulty meeting H_2S spec.

Influence of Impurities' and Table 16.6. Normally two or more sweetening processes are feasible. Further evaluation is made in the next step.

2. In the second step the following items are determined:

- *Claus gas composition*

  The Claus gas composition is an important factor in the selection, especially for gases containing H_2S and CO_2. A decision has to be made whether selective adsorption should be applied so as to keep the H_2S feed to Claus plant in the range that a proper flame is generated in the reaction furnace.

- *Solvent circulation rate*

  The solvent circulation rate is determined by the quantity of acid gas to be removed and the acid gas loading.

  Acid gas loading of the amine solvents is determined by the partial pressure and adsorption temperature. The maximum acid gas loading in practice is set by the approach to equilibrium and by acceptable corrosive levels. Table 16.6 shows the maximum practical acid gas loading.

- *Steam consumption*

  In most of the sweetening processes steam is necessary for regeneration. It is needed for:

  - raising the temperature of the solvent to regenerator bottom conditions (usually the boiling point)
  - desorption of acid gases from the solvent.
  - generating stripping steam, which will leave the regenerator top together with the acid gases.

  As energy saving is becoming a more and more important matter, it should be clear that the steam to solvent circulation rate is also a significant parameter. Note that physical solvents usually have a lower steam requirement per gallon of solvent circulated.

  The best means of determining how well the solution is stripped is by the H_2S in the stripped solution. Western Research has just developed an instrument for determining this.

  From the three items in the second step a preliminary selection and the most attractive process can be made, as solvent rate and steam consumption in general determine investment and operating cost. Of course solvent cost, royalties and existing plant facilities can influence the final choice.

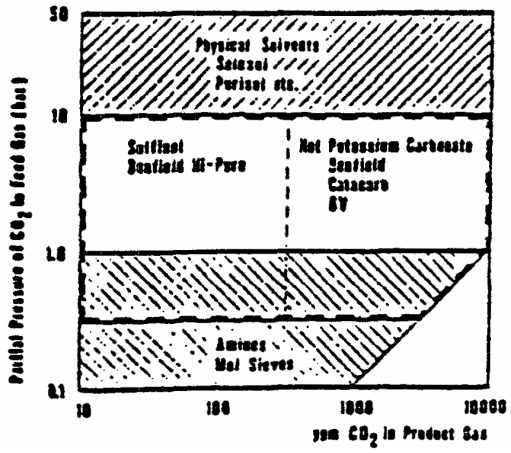


Figure 16.13: CO_2 Removal

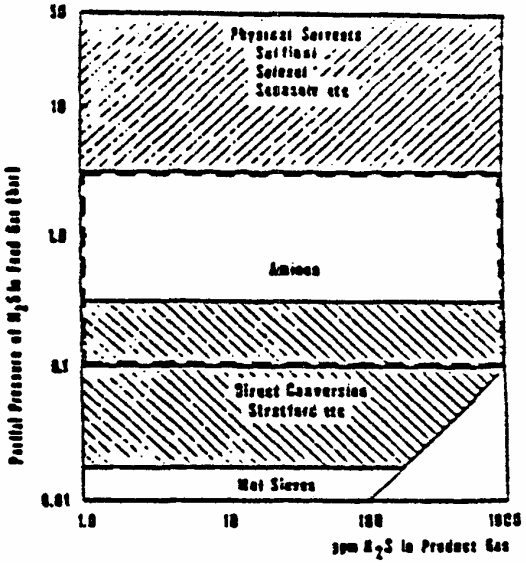


Figure 16.14: H_2S Removal

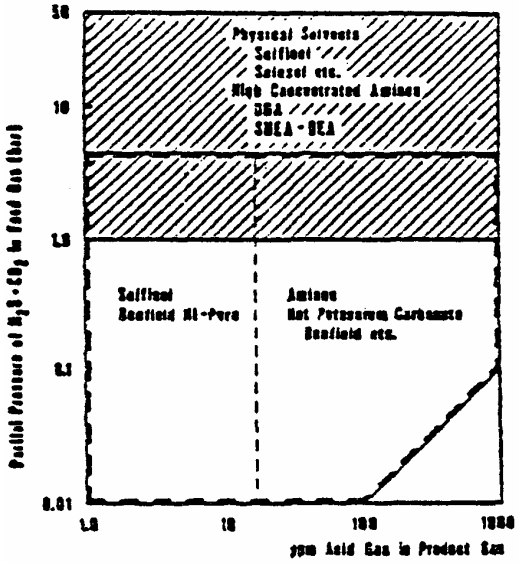


Figure 16.15: H_2S and CO_2 Removal

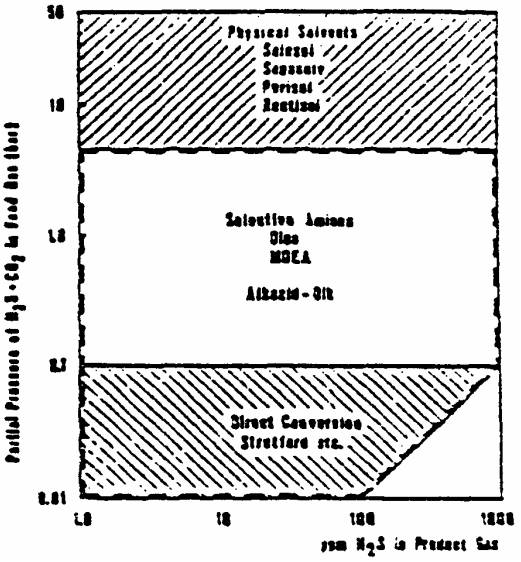


Figure 16.16: Selective H_2S Removal

Table 16.6: Maximum Practical Acid Gas Loading

Sweetening Agent or Process	Chemical Strength %	Solvent	Pick-up Moles Sweetening Agent/ Mole Acid Gas	Pick-up Acid Gas Scf/gal of Solution	Comments
MEA	20	H_2O	2 - 3		
MEA (with inhibitor)	30	H_2O	2 - 3		
DEA	25	H_2O	1.35 - 2		
SNPA - DEA	25	H_2O	0.9 - 1.3		
DEA (with inhibitor)	55	H_2O	1.35 - 2		

MDEA	50	H_2O	1.2 - 1.5		
DGA	50	H_2O	2 - 2.5		
Sulfinol	**	H_2O	----	6 - 17	** Sulfolane & DIPA 90%.
Selexol	----	----	----	up to 10.5	See notes for chemical.
Hot Potassium Carbonate	30	H_2O	2 - 3		Difficulty meeting H_2S spec.

16.5.2 Influence of Impurities

Impurities like COS, CS_2, mercaptans, HCN, NH_3 and heavy hydrocarbons will directly influence the process selection. For example:

COS and CS_2 as discussed previously, will react with MEA and will cause solvent losses. DEA, MDEA or DIPA are better choices in this case if an amine solvent is used.

COS and CS_2 are only partly hydrolysed in the activated hot potassium, so that some COS and CS_2 will remain in the product gas. In the Sulfinol process COS and CS_2 are satisfactorily removed. In the Selexol process COS and CS_2 are completely removed if the feed pressure is high. In the Stretford process COS is not converted and remains in the product gas.

In the gas from coal gasification plants many impurities are present like COS, CS_2, mercaptans, thiophenes, HCN, NH_3, aromatics, phenols, and some chloride and organic components. Depending on the gasification process and the type of coal which is used, these impurities will vary. HCN degrades most of the sweetening solvents. In Table 16.7 a survey is presented of the degree of absorption of the impurities, and of the solvent degradation by these impurities for the various processes. COS, CS_2 and HCN are sometimes hydrolysed over a catalyst bed to remove these impurities before the gas is treated in the sweetening plant.

Table 16.7: Absorption and Degradation of Impurities

	Degree of absorption of impurities							Degradation by impurities							Reclaimer required
I Chemical solvents	COS	CS_2	RSH	NH_3	HCN	Heavy HC's		COS	CS_2	RSH	NH_3	HCN	CO_2		
Amine															
MEA	M	M	S	C	C	No		Yes	Yes	No	No	Yes	No		Yes
DEA	M	M	S	C	C	No		No		No	No	Yes	No		No
SNEA-DEA	M	M	S/M	C	C	No		No	No	No	No	Yes	No		No
DIPA	S	U	S	C	C	No		No	No	No	No	Yes	No		No
MDEA	No	U	No	C	C	No		No	No	No	No	Yes	No		No
DGA	M/C	M/C	M			S		Yes	Yes	No			No		Yes
Hot Pot. Carb.															
Catacarb	M/C	M/C	No			No		No	No	No			No		No
Benfield	M/C	M/C	S	S	S	No		No	No	No	No	No	No		No
Giammarco Vetrocoke	M	M	S		C	No		No	No	No		Yes	No		No
Alkacid M	M	M	M		C	No		No	No	No		Yes	No		No
Alkacid DIK	M	M	M		C	No		No	No	No		Yes	No		No

C = almost complete removal.

M = approximately 50% removal.

S = slight removal.

U = unknown.

	Degree of absorption of impurities							Degradation by impurities								Reclaimer required
II Physical solvents	COS	CS_2	RSH	NH_3	HCN	Heavy HC's	H_2O	COS	CS_2	RSH	NH_3	HCN	CO_2	S		
Selexol	M/C	C	C	M/C	C	M/C	C	No	No	No	No	No	No			No
Purisol	C	C	C			M/C	C	No	No	No			No			No
Rectisol	C	C	C	M	C	M/C	C	No	No	No	No	Yes	No			
Sepasolv	M/C	C	C	M/C	C	M/C	C	No	No	No	No	No	No			No
Sulfinol	C	C	M/C	C	C	S	No	No	Yes	No	No	Yes	No			Yes
Amisol																
III Direct conversion																
Stretford	No	No	S	M	C	1)	No	No	No	No	No	Yes	No	Yes		2)
Takahax	No	No	S	M	C	1)	No	No	No	No	No	Yes	No	Yes		2)
Fugapol	No	No	No	No	No	No	No	No	No	No	No	Yes	No			
G.V. - Sulphur	S	S	S					No	No	No			No			
IV Dry bed																
Haines	C	C	C	C	C		C	No	No	No	No	No	No	No		
Molecular sieves	C	C	C	C	C		C	No	No	No	No	No	No			
Iron sponge	C	C	C	No	C			Yes	Yes	Yes	No	Yes	No			
Zinc oxide	C	C	C	No	C			Yes	Yes	Yes	No	Yes	No			

1) gives sulphur contamination.

2) effluent treatment.

16.6 H_2S Scavenging Processes

For single or multiple moderately sour gas wells where the total production is small and the location remote from a gas processing plant, a conventional sweetening plant may be uneconomical due to the large investment required. Moreover, an amine plant will remove CO_2 as well as H_2S while the purpose is only H_2S removal. In such cases, the option of using other cheaper sweetening processes, such as batch operated chemical H_2S scavengers, must be considered. Concentrations of 500 ppm or less should be considered for these types of processes.

The table below, Table 16.8, outlines the various processes that have been investigated by one large company. Also, in the Table 16.9 qualitative rating of the major processes is given.

Table 16.8: H_2S Scavenging Processes

Formaldehyde-Methanol Based
Scavinox
Prohib 196
Champion Exp Cortron JS-49-4
Baker Magnatreat M118W (formerly Dichem E118W)
Techniscav

	Couger H-925
Caustic Based	
	NL Chemicals Sulfa Check
	Champion Gas Treat 102
Chlorine Dioxide Based	
	Surflo 2341
Metallic Oxide Based	
	Slurrisweet (now called Ecotreat)
	Chemsweet
	Iron Sponge
	ICI Zinc Oxide Dry Bed

Table 16.9: Qualitative Process Comparison

	Scavinox	Sulfa-Check	Chemsweet	Slurrisweet	Iron Sponge	Di Chem
Capital Cost	4	5	0	1	2	3
Operating Cost	5	2	3	4	0	1
Proven Process	4	4	4	3	5	4
Winterization Rqmt.	5	0	0	0	0	5
Ease of Operation	4	4	4	2	4	4
Operator Acceptance	0	5	5	3	2	0

Scale: 0 = worst to 5 = best

16.7 Problems

Example Problem 16.1

A sour gas of the composition given in the table is to be treated by using a split stream MEA solution of a 25% by weight strength. The flow of sour gas to the contactor is 150 MMSCFD. The total solution circulation is such that 2.0 moles of amine are circulated/mole of acid gas. Approximately 90% of the circulation is through the bulk removal section. This solution is regenerated such that the H_2S concentration in the semi lean amine that has been regenerated is 100 grains per gallon. The clean-up amine is to be regenerated to 10 grains of H_2S/U.S. gallon.

Gas Composition	
Compound	**Mole %**
H_2S	3.5
CO_2	5.5
C_1	79.8
C_2	5.2
C_3	2.4
iC_4	0.8
nC_4	1.3
C_5	1.5

Determine the steam required in lbs/hr by the plant for the regeneration of the solution.

Example Problem 16.2

An existing plant is treating 100 MMSCFD of inlet gas of the composition as given in column 1 below. A conventional mono-ethanol amine system has been used to treat the gas (3 moles of MEA per mole of acid gas with a 20% solution of MEA in water). The sales gas had only to meet Trans Canada specification for H_2S content.

A field is discovered nearby that has gas of the composition given by column 2 in the table. The maximum flow of raw gas from this field is 40 MMSCFD.

It has been proposed that instead of building a new sweetening plant that the gas be brought into the present plant and the sweetening solution changed to diethanol amine. The licensors of the SNPA processes recommend that the maximum solution rates and strength of the patent be used. Also that the loading of each gas in amine be the minimum but still in the patent range.

Determine whether the existing plant can handle the total raw gas as now contemplated.

1. The amine contactor, exchangers and pumps are sized by liquid loading.
2. The amine still and its equipment are sized from steam required for regeneration.

Comment on the result of your investigation.

Raw Gas Flow $\frac{\#1}{100}$ $\frac{\#2}{40}$

MMSCFD Comp.	Mole %	Mole %
H_2S	6.0	15.0
CO_2	1.0	1.0
C_1	82.0	81.0
C_2	5.0	2.0
C_3	1.7	1.0
iC_4	0.8	--
nC_4	1.5	--
C_5^+	2.0	--

Example Problem 16.3

A gas of the following composition has to be sweetened. Tell the process you would use and why.

N_2	0.50
H_2S	3.25
CO_2	15.35
C_1	78.70
C_2	2.10
C_3^+	0.10

Draw a process flow sheet of the process you choose and estimate the cost of 100 MMSCFD.

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17.0 Dewpoint Control (Hydrocarbon and Water)

- A detailed review of each of the processes used for dewpoint control, including design criteria, operating problems, etc.

17.1 What Are We Trying to Do?

You will recall at the beginning of the book we mentioned specifications for the water content of pipeline gas (usually 4 lb of H_2O/MMSCF which is equivalent to a dewpoint of about 15ºF at 900 psia) and for the hydrocarbon dewpoint of pipeline gas (usually 15ºF at 800 psia). The meeting of these specifications can often be accomplished by a single processing scheme. This section reviews these schemes and then the processing schemes that are applicable to the meeting of only one of the specifications. Table 1 below lists the schemes discussed and their applicability.

Table 17.1: Comparison of Dewpoint Control Schemes

Process Scheme	For H_2O Content Specification	For Hydrocarbon Dewpoint Specification
Chilling with Glycol Addition	✓	✓
Short Cycle Adsorption	✓	✓
Glycol Dehydration	✓	
Long Cycle Adsorption	✓	
Oil Absorption		✓
JT Throttling with Methanol	✓	✓

17.2 The Chilling Process

17.2.1 Process Description

This is quite a simple process. It is one of the most common processes used in Western Canada. Such plants as the East Calgary, Rimbey, Edson, Sylvan Lake Plants, etc., use the process. The process basically consists of cooling the gas to approximately 5ºF (.3ºC) below the required dewpoint by series of heat exchangers. The final cooling is obtained by a refrigerated chiller generally using propane as the refrigerant. Ethylene glycol is injected into the gas stream at the beginning of the process. The glycol combines with the water that is condensed and prevents the forming of hydrates or ice by this water. The glycol water solution is separated from the hydrocarbon that also condenses and the glycol is suitably regenerated. Figure 17.1 is a process flow sheet for the process.

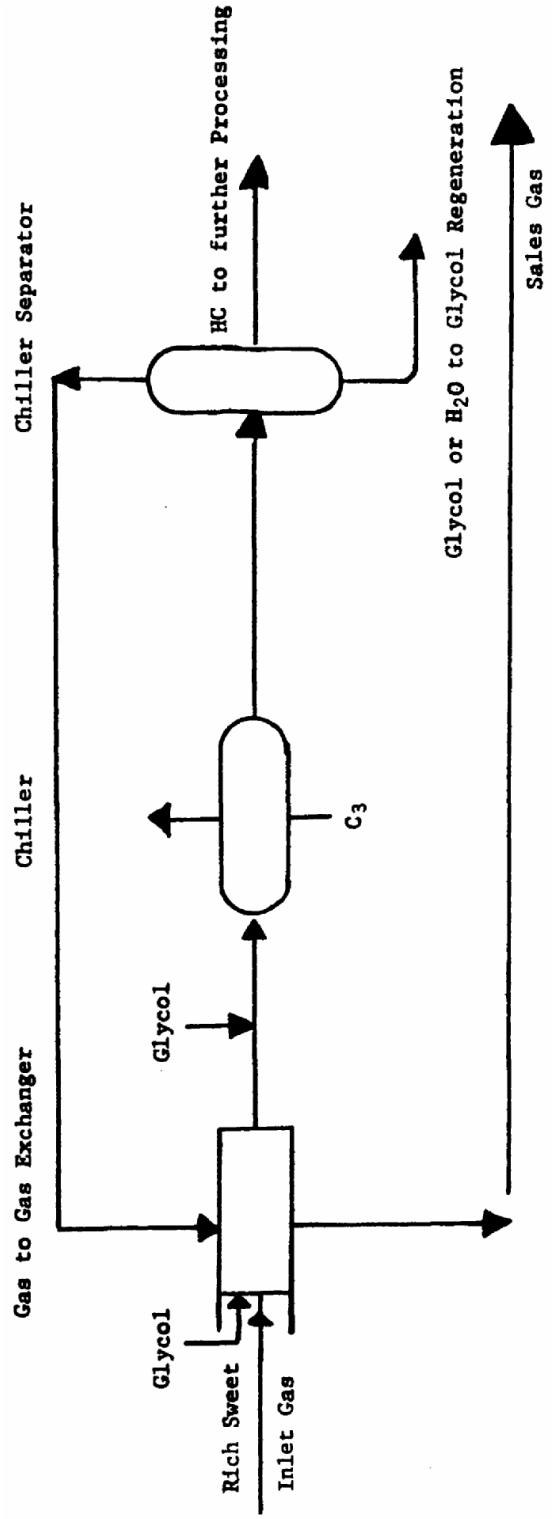


Figure 17.1: The Chilling Process

17.2.2 Design Criteria

The principles used in design of this process are quite straight forward. Primarily the requirements are the proper sizing of the heat exchangers. This includes determining the total heat to be removed. This, of course, includes the heat of condensation of liquid hydrocarbon.

One of the keys in the successful operation of such a unit is the proper selection of a glycol to inject and how much to inject.

Normally mono-ethylene glycol is used as the injection glycol as it separates more easily at law temperatures and does not absorb as much hydrocarbon as the higher molecular weight glycols. The actual concentration of mono-ethylene glycol to be used can be calculated by the Hammerschmidt equation as given in Chapter 5.

Equation 17-1

$$d = \frac{4000W}{100M - MW}$$

or in metric

Equation 17-2

$$d = \frac{2222W}{100M - MW}$$

where:

d = number of degrees of depression required to prevent freezing or hydrate formation.

M = molecular wt. of glycol used.

W = weight % of final glycol concentration.

Now actual glycol circulation is calculated by a water balance assuming an inlet concentration and then calculating "W". For safety we add 100% to this calculation to determine the glycol circulation. Example problem No. 3 in Chapter 5 shows the application of this equation.

The glycol concentration is usually about 80 to 85 weight % when being injected and the concentration is reduced to between 50 and 60 weight %.

Glycol distribution is another problem. Great care should be taken with the distribution of glycol into the gas. It is particularly important that the spray nozzle that is used be picked bearing in mind the operating pressure of the system. Also the glycol should make good contact with the gas throughout the whole gas flow.

17.2.3 Operations

17.2.3.1 Glycol Losses

These should run about 3/4 to 1 lb/MMSCF in a properly designed system. This is greatly affected by the separation of the glycol water solution from the condensed hydrocarbon layer. This can be improved by warming the two materials before the final separation. Often stages of separation can be profitably installed.

17.2.3.2 Glycol Foaming

Besides good hydrocarbon glycol separation, a good filtering system is desirable for keeping the solution clean.

A good reference on the design of glycol injection system is Robirda & Marten.

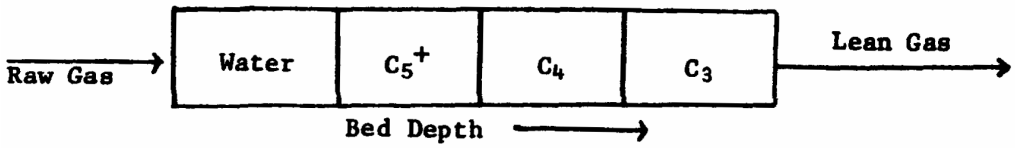
17.3 The Adsorption Process

In this process a solid desiccant is used to absorb the heavier hydrocarbons while letting the lighter hydrocarbons through. This is also good for drying (removing water) gas.

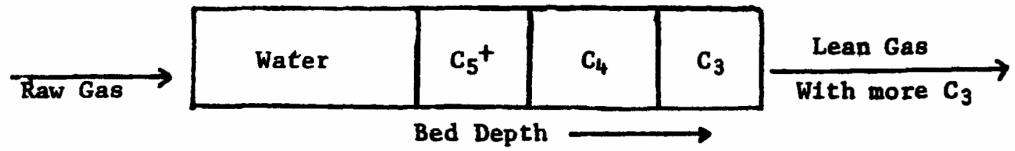
A short review of adsorption theory follows.

Initially the wet (rich) gas contacts the bed. The water then C_5^+. C_4 and C_3 are adsorbed in order.

Initially:



Then later the following is a picture of the bed:



As can be seen the water is using up more of the bed and less hydrocarbon is being adsorbed.

17.3.1 Adsorbents

These are charcoal, silica gels, alumina and synthetic zeolite materials with a large porous surface area.

For example silica gel has an internal surface of 800 sq. meters per gram of material.

The mechanism is that molecules enter pores of adsorbent and condense, particularly if they are polar. Also pores act as capillaries and prevent the revapourization of the material. Adsorbent pores can be sized so as to allow only certain molecules to enter.

To recover hydrocarbon short cycles are used so that all the bed is not used up in adsorbing water. Heavy hydrocarbon displaces lighter hydrocarbon and water displaces the heavy hydrocarbon.

17.3.2 Process Description

The process is quite simple. The rich gas enters one tower filled with desiccant. The desiccant adsorbs the various products until the cycle is complete. Then the towers are switched and heat applied to drive the adsorbed components off which are then sent to a condenser and a separator and the liquid products are recovered. The flow sheet (Figure 17.2) shows the major pieces of equipment.

In some variations a 3 bed cycle is used one for adsorbing, one for regenerating and one for cooling.

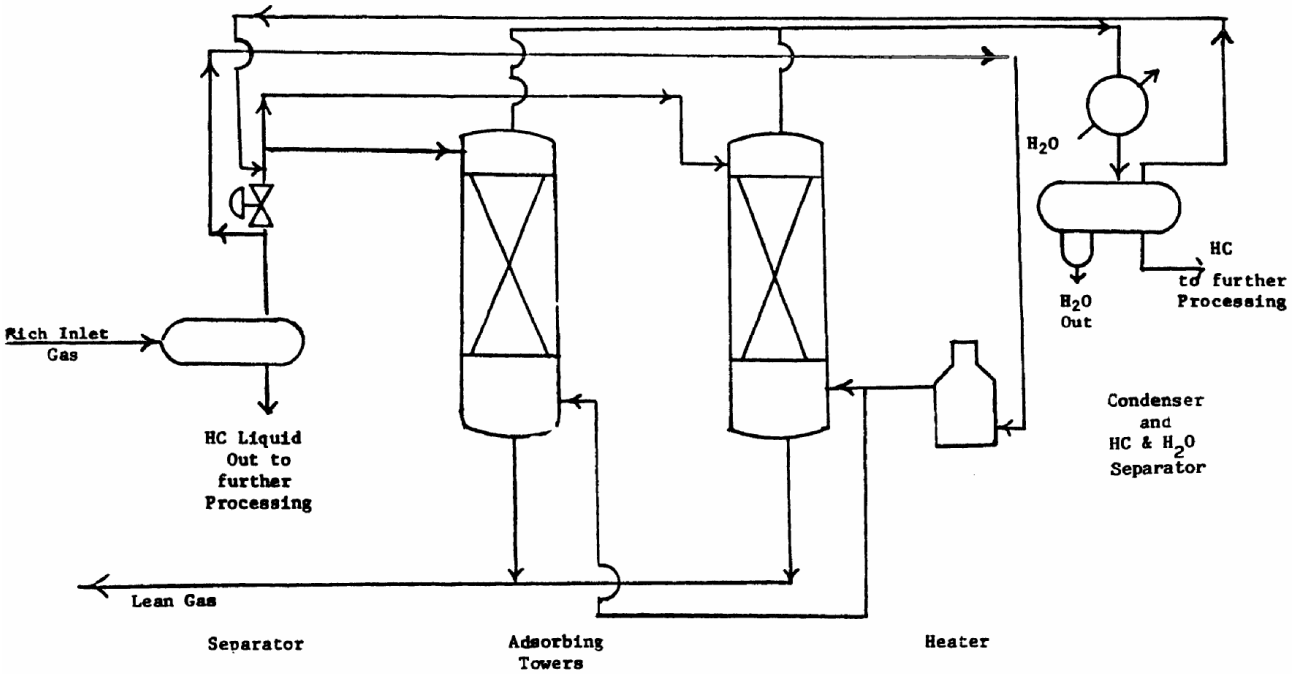


Figure 17.2: Adsorbing Process

17.3.3 Design Considerations

There are several design conditions. These include:

1. Cycle time - This is set primarily by the length of time it takes to regenerate a bed. That is, the length of time it takes to heat up a bed to a regenerating temperature, and if necessary, cooling some to get it ready for the adsorbing cycle.
2. Gas Composition - This probably is the most important variable associated with a solid bed unit and that place where most of the problems arise, particularly when trying to put in a limit on a new gas field. This problem is discussed in Section 12, “Producing of Gas Well”, under “Sampling”. The gas composition, the recovery of products and the cycle time sets the amount of adsorbent required. As C_3 and C_4 are normally recovered early in the cycle, a high C_3 and C_4 recovery requires a large bed.
3. Pressure - The lower the pressure, the greater the adsorption efficiency.
4. Temperature - Lower temperature improves operating efficiency.
5. Contaminants - There are several contaminants. Heavy ends and compressor lube oils can cover over the adsorbing pores and cause the desiccant to Lose its efficiency. Similarly glycols can do the same.

   Another material that is bad for the silica gel-type desiccant is amine or other ammonia compounds.
6. Typical Adsorbents - The primary adsorbent used is silica gel or the variation of the same called mobil beads.

   Another adsorbent is activated carbon which has come back into use recently on pipeline

gas. However, considerable problems have arisen from these plants - primarily due to the high reactivation temperatures (680ºF to 700ºF).

7. Special Data - Some typical design criteria when getting a solid bed hydrocarbon unit are as follows:

   a) Minimum cycle time depends a great deal on time required to regenerate beds and to cool them for re-use. Twenty to thirty minutes is usually best.

   b) Amount of Desiccant. This is particularly affected by gas analysis and the type of desiccant. Data determining this for Mobil bead type of adsorbents (silica gel) is given in the article by C.L. Humphries, "Hydrocarbon Processing" , and for silica gel by J.M Campbell.

The total amount of dessicant required is found by calculating the lbs of dessicant required to absorb each component taking into account preloading of the bed during the cooling cycle. For each component there is a lb of component absorbed per 100 lbs of adsorbent. A typical set of values for silica gel is shown on the figure below.

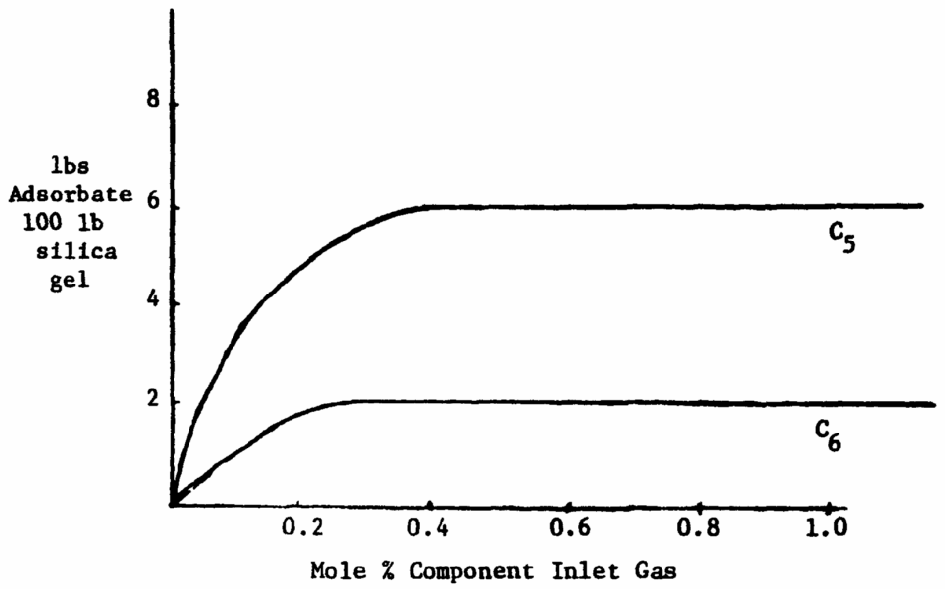


Figure 17.3: Adsorbing of C_5 & C_6

This curve is based on 45 ft/minute flow through bed (This is normal design superficial flow rate). Also please note that the adsorptive capacity for each component is affected by the presence of the other components. This particularly applies to C_3 and C_4 adsorption. For further data on component adsorption please see Chapter 7, “Mass Transfer”.

17.3.4 Operations

This process has had many operating problems. One of worst has been that the beds do not adsorb as much as they were for. This has been primarily caused by feeding the unit of gas different analysis than expected. Particularly the presence of heavy ends causes a lot of trouble.

Also it is necessary to make sure that there is good regeneration of the bed. That is, the whole bed is raised to the design regeneration temperature.

17.4 Dehydration Processes Only

The previous two processes discussed have been for both water and hydrocarbon dewpoint control. In this part we will consider two processes that are used for the removal of water only. These are the use of a glycol for dehydration and the use of a solid bed desiccant for dehydration. These processes have been referred to in the field processing of natural gas section.

It is desirable to review the properties of the various glycols and desiccant before describing the processes. These are given in Table 17.2.

Table 17.2: Properties of Glycols & Solid Desiccants

A. Glycols Properties			
Name	**Ethylene Glycol**	**Diethylene Glycol**	**Triethylene Glycol**
Abbreviation	EG	DEG	TEG
Formula	HOC_2H_4OH	$HOC_2H_4O\text{-}C_2H_4OH$	$HOC_2H_4O\ C_2H_4O\text{-}C_2H_4OH$
Molecular Wt.	62.1	106.1	150.2
Boiling Point ºF	387	473	546
Vapour Pressure at 77ºF mm of Hg	0.12	< 0.01	< 0.01
Gravity 77ºF	1.110	1.113	1.119
Freezing Point ºF	8	17	19
Flash Point	240	280	320

B. Solid Desiccant Properties					
	Bauxite	**Alumina 151**	**Mobil Beads**	**Davison Silica Gel**	**Molecular Sieve**
Main Chemical Constituent	Al_2O_3 Plus	Al_2O_3	SiO_2	SiO_2	Synthetic Zeolites
Density lb/cu. ft.	50	52	49	45	43
Absorptive ** **capacity now at 40% RH, percent**	6.0	14	22.7	24.85	24.0
Regeneration Temp. ºF	350	350	300 - 500	250 - 500	300 - 600
Relative Pressure Drop	High	Low	Low	High	Average
Particle Shape	Granular	Sphere	Elongated Sphere	Granular	Pellets

** See Figure 17.5 on relative capacity of desiccants with time.

17.4.1 Glycol Dehydration

17.4.1.1 Process Description

The process is quite simple. Glycol is circulated through a tower in which the wet gas is passed through. The glycol absorbs the water vapour from the gas. The glycol is then reduced in pressure as much as possible and then heated in a reboiler to regenerate it - Figure 17.4 shows the process.

17.4.1.2 Design Data

a) Major Equipment - The major equipment consists of a contactor containing about four trays, a regenerator with a fired reboiler in which the heat absorption rate is generally less than 6,000 BTU/hr.sq.ft. The regenerator usually has about 4 to 6 trays, or packed section with about 12 ft. of packing. This, of course, varies with the dewpoint suppression required.

Generally a filter is used to remove iron sulphide and other particles that cause foaming.

b) Circulation Rate - A glycol circulation rate of 3 to 5 gallons per lb. of water is a good design number.

c) Regeneration Temperatures - For DEG: 325-350ºF, for TEG: 400ºF (Max.). Primarily limited by possibility of decomposition.

d) Normal Lean Solution Concentrations – DEG: 95-96% by wt., TEG: 98-99% by wt.

e) Dewpoint Suppression - The amount of dewpoint suppression is dependent on the quality of the glycol to be circulated. A good reference for determining this is the Dow Gas Conditioning Fact Book.

For example, using a 99.7% TEG material and the outgoing gas is in equilibrium with the glycol, a dewpoint of -20ºF is possible if the inlet, TEG is at 80ºF or less. Here is some data on diethylene and triethylene.

Table 17.3: Comparison of DEG & TEG Performance

Item	Diethylene Glycol	Triethylene Glycol
Normal strength for dehydration	95%	98.5%
Regeneration Temperature	330ºF (165)	360ºF (182)
Normal dewpoint suppressions	65ºF (36)	90ºF (50)
Dewpoint suppression with gas stripping	110ºF (61)	140ºF (78)

17.4.1.3 Operating Problems and Operating Data

a) Glycol Losses - The primary operating cost of these units is glycol loss. Good inlet separators and filtering should reduce foaming tendency; also sometimes an antifoam agent is added.

Normally glycol losses should be 3/4 to about 1 lb. per MMSCF for sweet gas; 3 to 4 lb. per MMSCF with sour gas.

b) Corrosion - Glycol may become corrosive, particularly when treating sour gas. Also, air should be kept out of the system. Air causes the breakdown on glycols to organic acids. Various inhibitors are sometimes used.

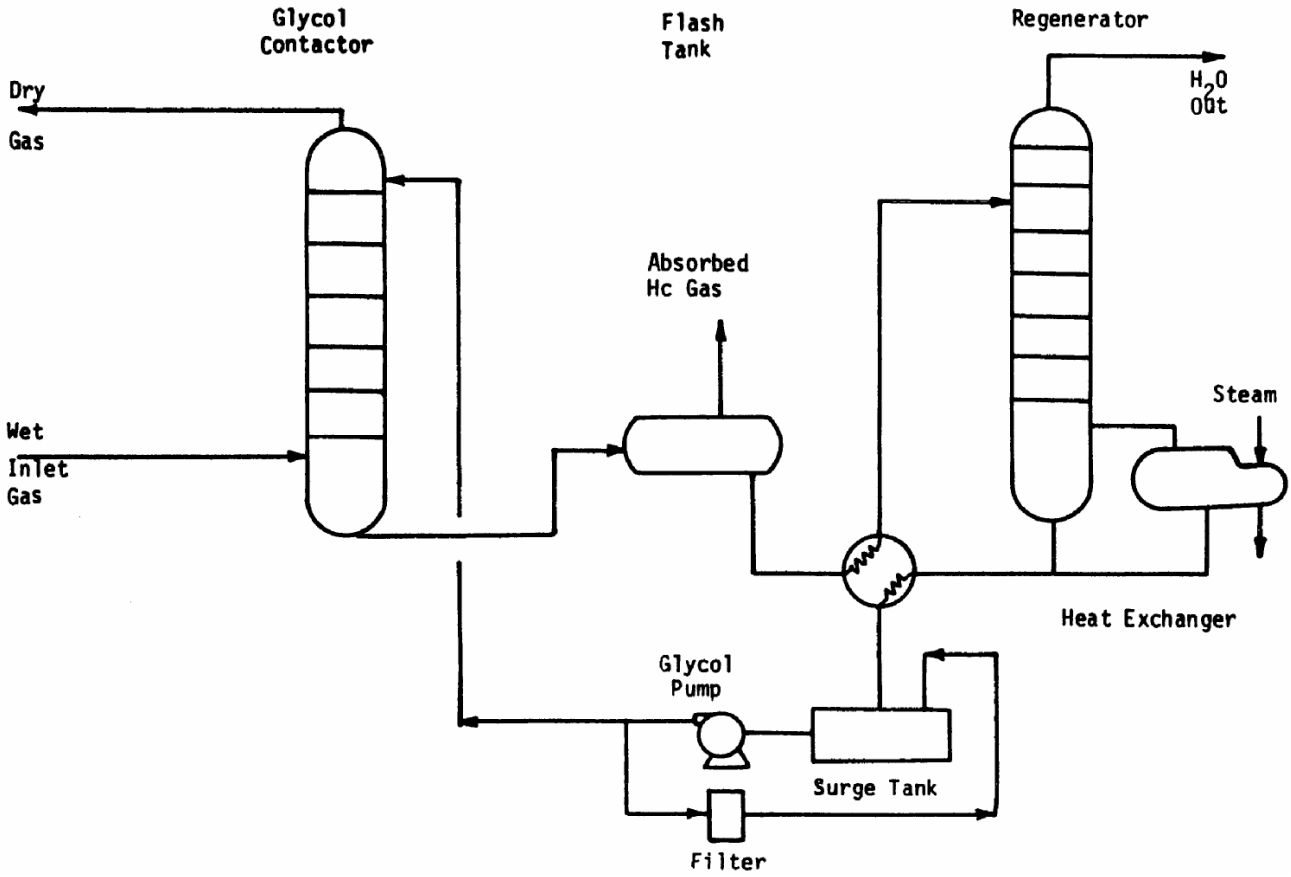


Figure 17.4: Glycol Dehydration

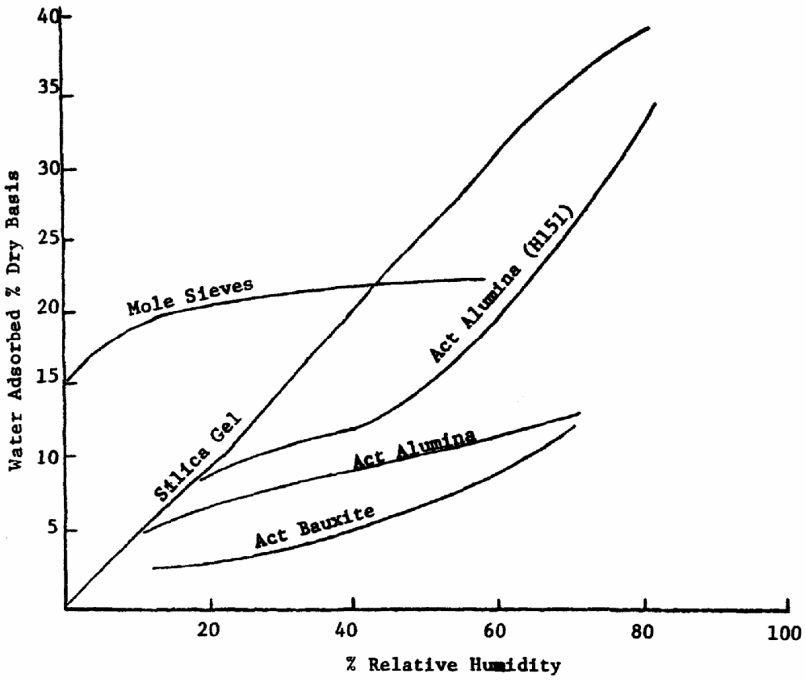


Figure 17.5: Equilibrium Capacity Solid Adsorbents

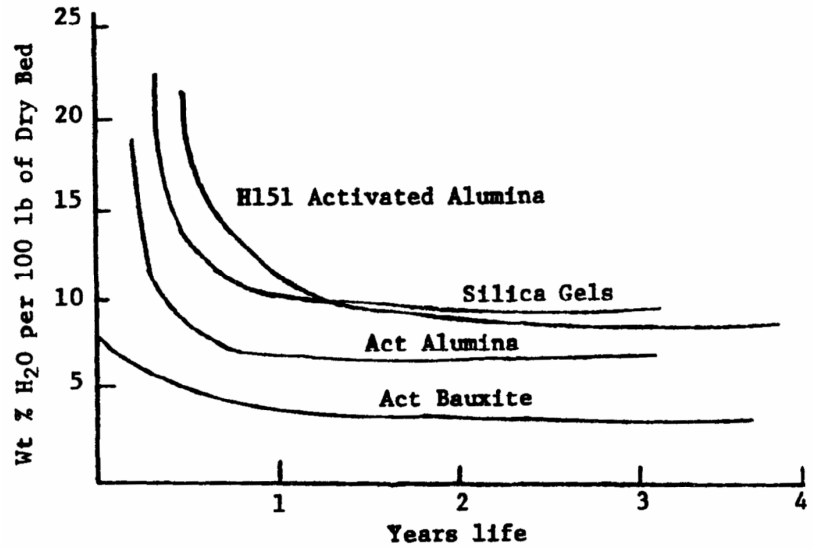


Figure 17.6: Commercial Operation of Solid Desiccants Manufacturers Data

17.4.2 Long Cycle Adsorption

17.4.2.1 Solid Desiccant Drying

This process uses several different adsorbents. These are bauxite, alumina, silica gels and synthetic zeolites. These materials have different degrees of adsorptivity and thus their selection is based on economic considerations.

Note: Physical properties are given in Table 17.2.

17.4.2.2 What Happens in Adsorption Tower?

As wet gas enters the tower the water vapour is condensed and removed in the first part of the tower. Then as time passes the bed becomes progressively saturated until eventually the whole bed becomes saturated. Figure 17.7 shows what happens to the dewpoint with time.

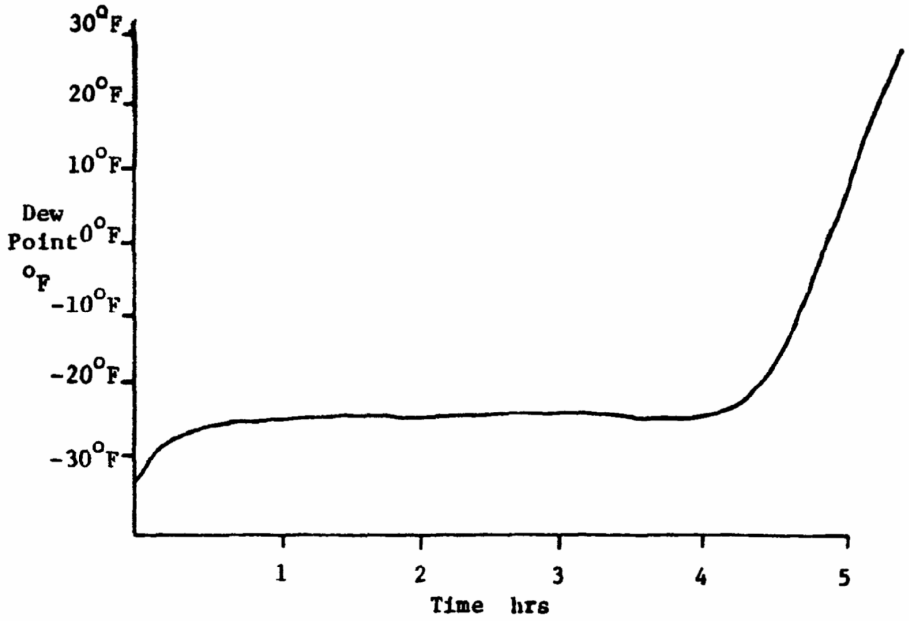


Figure 17.7: Dewpoint Values Out of a Desiccant Bed

17.4.2.3 Regenerating of an Adsorption

The regeneration of the adsorption tower is normally accomplished by heating with a hot gas. The time-temperature curve observed when accomplishing this is shown in Figure 17.8.

Initially the bed is heated to vapourizing temperature of water vapour; then a constant temperature is maintained while all the water is driven off. Then the temperature rises again while heavy hydrocarbons are driven off. If there is a large concentration of one compound, a plateau occurs; if not, the temperature continues to rise. Regeneration can stop when the bed outlet temperature reads 325 to 350ºF.

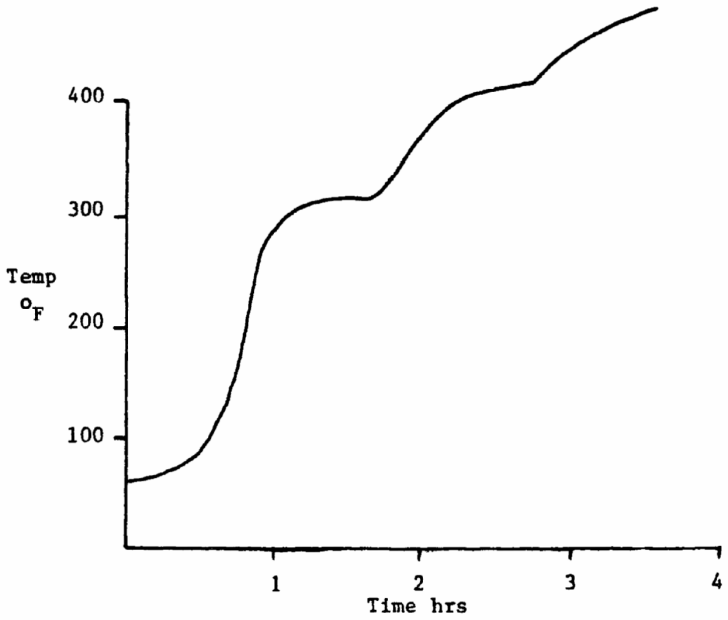


Figure 17.8: Regeneration of a Desiccant

17.4.2.4 Process Description

The wet gas enters tower at the top and passes out the bottom as a dry gas. A slip stream of wet gas is taken through a heater, then passed through the dryer, being reactivated in a CO—current direction and the bed heated. When the bed has been heated to the desired reactivation temperature then the heater is shut down and the bed is cooled. The gas leaving the tower, being regenerated, is cooled in a condenser and the water condensed out. The gas is sent back to the wet gas stream. The process flow is similar to that shown in Figure 17.2.

With a 3 dehydrator system, a slightly different system, in which one tower is regenerated and one tower is cooled, is normally used.

There are variations in the flow in that a split flaw design can be used but it is not too successful.

17.4.2.5 Design Data

Size contacting equipment by finding how much desiccant is required. Some good figures to use are:

Activated bauxites	4%
Activated Alumina	6%
Activated Alumina H-151	6%-8%
Silica Gels	6%-8%

Thus, if we know the amount of water to be removed we can calculate the amount of desiccant to use. Then to find diameter and depth of bed, pick an allowable velocity through a tower. This is normally taken as 30 to 60 ft/min (Superficial).

17.4.2.6 Other Criteria

- Regeneration gas flow 5%-10% of main flow.
- Regeneration heat requirements: Calculate both the heat loss from the unit and latent heat of evaporation of water.
- Must allow in cycle time enough time for heating, ie. to reach design regeneration temperature on exit of bed and time for cooling.

17.4.2.7 Operating Problems

One of the special problems with solid bed units is that they absorb H_2S which is driven off during part of the cycle. This can cause a high H_2S reading in outlet gas.

When using bauxite it is also necessary to watch that it is iron free, as Fe_20_3 reacts with H_2S to form iron sulphide. Also, if air is present $2H_2S + 0_2 \Rightarrow$ can form free sulphur. Bauxite is the normal catalyst in this reaction.

Oil can contaminate desiccant very early by filling pores with non-regenerable compounds. However, plants downstream from oil absorption plants have worked satisfactorily.

Amine is also a contaminant for silica gel.

Note: Proper screens are required to hold desiccant. Several problems have developed because of poor screens.

17.5 The Absorption Process

This process is often used when it is desirable to recover more than condensate from the gas. i.e. lpg's C_3 and C_4. Quite often it is used in conjunction with a chilling (refrigeration) process as at Rimbey or Edson, or directly, as at Pincher Creek or McMahon Plant.

The unit operation of absorption has been discussed in some detail in Chapter 7, “Mass Transfer”.

17.5.1 Process Description

The flow sheet (Figure 17.9) shows the process used. Primarily the process is contacting the gas with a lean oil which absorbs the heavier components. The rich oil is then reduced in pressure and heated to strip out the absorbed product and the oil recirculated. The lpg's are fractionated to recover them. The process is described in more detail In the LPG Recovery, Chapter 19 and in Chapters 18 and 7.

17.5.2 Design Criteria

There are many variables in the absorption process. One of the first items to consider is the oil molecular weight. The absorption Fraction $A = \frac{L}{KV}$ is based on mole weight.

Thus the lighter the oil the more moles can be circulated and thus better absorption can occur. This has to be balanced by absorption oil losses from vapourization. The following table shows a rough range of ‘Oil MW versus Temperature’.

Table 17.4: Absorption Mole Wts.

Temperature ºF	Mole Wt.
+80	180 - 200
0	140 - 160
-30	100 - 120

For example, the Rimbey plant, which goes after high C_3 recoveries, use low mole wt. oil.

Another characteristic of absorption oil should be that it is highly paraffinic. Cracked or aromatic stock should not be used as it causes foaming.

The determination of the number of trays in an absorber has been covered in the Mass Transfer chapter.

Another noteworthy part of the process is the presaturation of oil. This is done by contacting the oil with a lean gas high in methane to reduce the methane absorbed in the main absorber. This, of course, also reduces the heat of reaction in the main absorber and allows the other components to be absorbed at a low temperature.

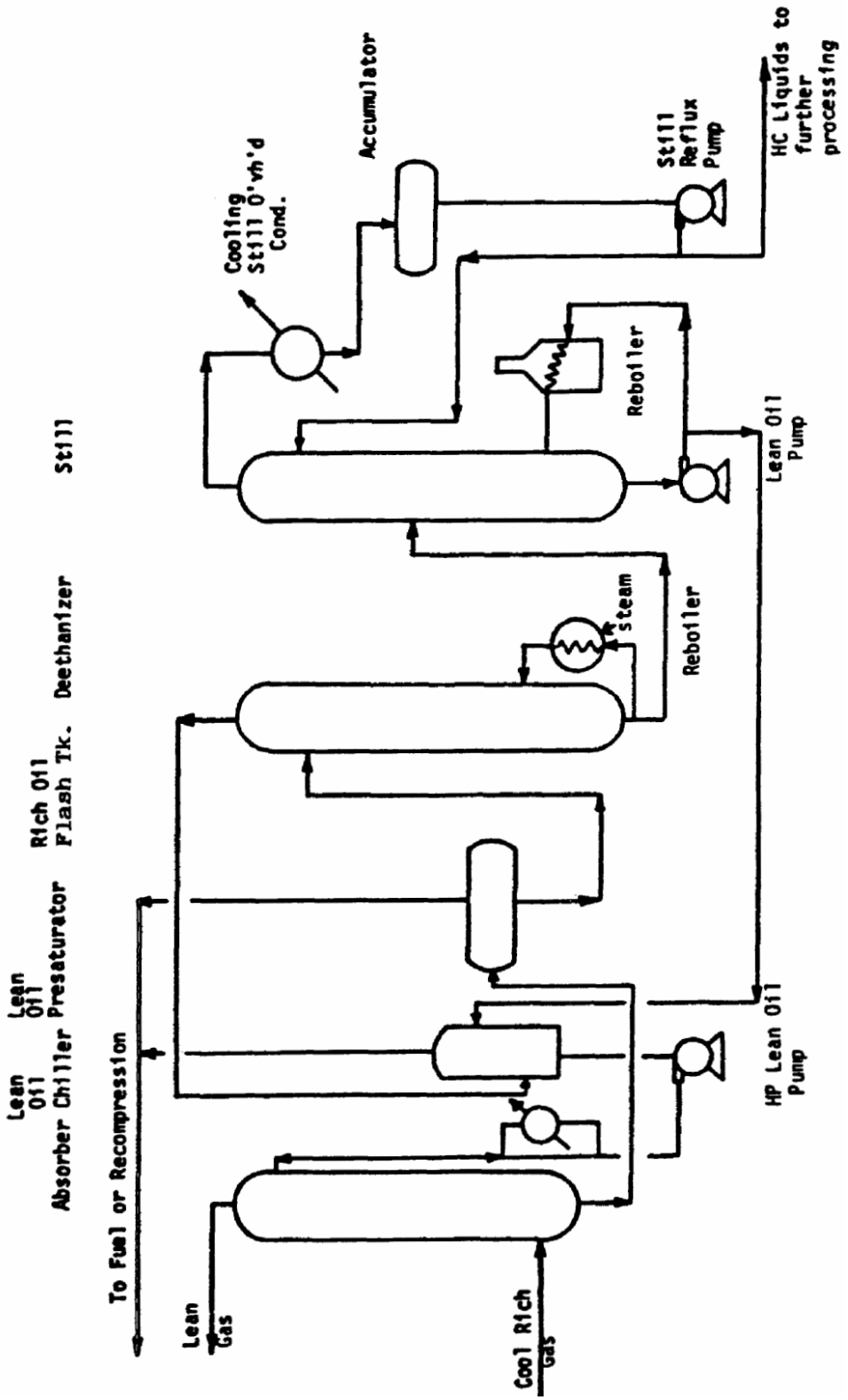


Figure 17.9: The Absorption Process

17.5.3 Operations

There are few operating problems in the absorption process. Initially absorption oil losses are high, but eventually they can be controlled and often the system does not need outside absorption oil.

References

1. Robirds, K.D., and J.C. Martin, Oil & Gas Journal, April 30, 1962, p. 85.
2. Humphries, C.L., Hydrocarbon Processing 45, No. 12, P. 88, (Dec.1966).
3. Campbell,J.M., Hydrocarbon Processing 42, No. 12, P. 89, (1, Dec. 1963).
4. Dow Chemical Gas Conditioning Fact Book, Dow Chemical Company, Midland, Mich., 1962.

18.0 Fractionation

- A short review of the processes used.
- A discussion on condensate stabilization.

18.1 Introduction

In the discussion on mass transfer (Chapter 7), we discussed distillation or fractionation as the separation of Components by the application of heat to the more volatile components and we saw that this process consisted of a series of vaporizations and condensations. We also discussed methods of calculating the height and diameter of towers used for this process.

Under hydrocarbon dewpoint control we saw that the hydrocarbon product was recovered by a form of fractionation. The product produced from such an operation still contains several commercial products. These include C_3, C_4, C_5 and others.

In this section, and in further sections, we will review the application of fractionation to the separation of these products. Some principles should be reviewed. These include:

a) A tower normally separates only two pure components. Thus if you have to separate three components, two towers are necessary. Side-draw towers can sometimes be used, but this is not often the case.

b) The difficulty of separation is measured roughly by the difference in boiling points or, more accurately, by the difference in "vapour-liquid "K" values. This is called the relative volatility α and is represented by the equation:

Equation 18-1

$$\alpha = \frac{K_{\ell k}}{K_{hk}}$$

where:

$K_{\ell k}$ = the equilibrium constant for the light key component.

K_{hk} = the equilibrium constant for the heavy key component.

A typical tower is as follows:

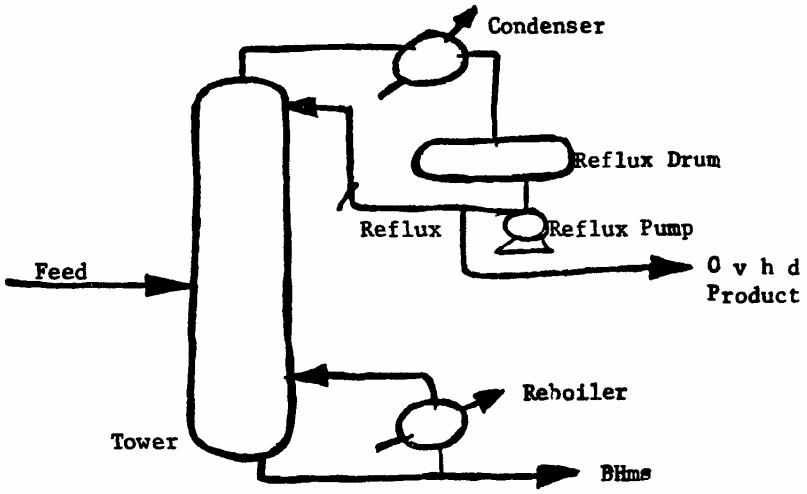


Figure 18.1: Typical Fractionating Tower

18.2 Tower Sizing

The sizing of the fractionation towers was covered in detail in the Mass Transfer Chapter. It will be briefly reviewed here.

18.2.1 Tower Diameter

The actual tower diameter is determined by the flow of fluids in the tower (generally the vapour flow) and by the type of internals in the tower. Generally it is best to have a company that specializes in tower internals size the tower. However, the approximate diameter can be determined by calculating the allowable velocity from the formula:

Equation 18-2

$$G = C\sqrt{\rho v(\rho L - \rho v)}$$

where:

ρv = vapour density lbs/cu.ft.

ρL = liquid density lbs/cu.ft.

The constant "C" is obtained from Figure 18.2 and is dependent on the type of tray internals and the tray spacing.

Note: Generally 24" spacing is used, although sometimes 18" spacing is used.

18.2.2 Tower Height

The tower height is determined by the number of trays, their spacing, the spacing allowable for disengaging at the top of the tower and the amount of space allowed for in the bottom of the tower for handling of the liquid out of the tower. One way of determining the number of trays is as follows:

1. Determine the minimum of trays (at infinite reflux).
2. Determine the minimum reflux (at infinite number of trays).

3. Assume an economic reflux usually 1.25 to 1.3 times the minimum. With the high cost of energy, probably a lower value around 1.10 to 1.20 is better now.
4. Calculate the actual number of theoretical trays using the nomograph shown in Figure 3.

The minimum number of trays can be determined from Fenske' s formula:

Equation 18-3

$$(\alpha_{LK})^{Sm} = \frac{(X_{LKD})(X_{HKW})}{(X_{LKW})(X_{HKD})}$$

where:

α = the relative volability of the separation.

Sm = minimum number of theoretical trays for the separation.

X_{LKD} = moles of light key in the overhead product/100 moles of feed.

X_{HKW} = moles of heavy key in the bottoms product/100 moles of feed.

X_{HKD} = moles of heavy key in the overhead product/100 moles of feed.

X_{LKW} = moles of light key in the bottoms product/100 moles of feed.

The minimum reflux can be determined by the 'Underwood" method using the following equations:

Equation 18-4

$$\frac{\alpha_1 X_{1F}}{\alpha_1 - \theta} + \frac{\alpha_2 X_{2F}}{\alpha_2 - \theta} + \ldots\ldots\ldots\ldots = 1 - q$$

Equation 18-5

$$\frac{\alpha_1 X_{1D}}{\alpha_1 - \theta} + \frac{\alpha_2 X_{2D}}{\alpha_2 - \theta} + \ldots\ldots\ldots\ldots = R_{\min} + 1$$

Equation 18-6

$$q = \frac{H_V - H_F}{H_V - H_L}$$

where:

α = relative volatility, ie. ratio of K for compound to K of heavy key.

X_F = mole fraction in feed.

X_D = mole fraction in overhead product.

R_{min} = minimum reflux.

H_V = enthalpy of a saturated vapour of the composition of the feed.

H_L = enthalpy of a saturated liquid of the composition of the feed.

H_F = enthalpy of the feed at its condition.

θ = a function obtained by using Figure 30 in Section 7 as first approximation.

Subscript

1 moles or “α” of component 1

2 moles or “α” of component 2

n moles or “α” of component n

Please note that efficiency has to be taken into account when the actual number of trays are determined. This efficiency can be calculated as given in the Mass Transfer Chapter or as an approximation for the separations considered in this chapter. Seventy-five percent can be used.

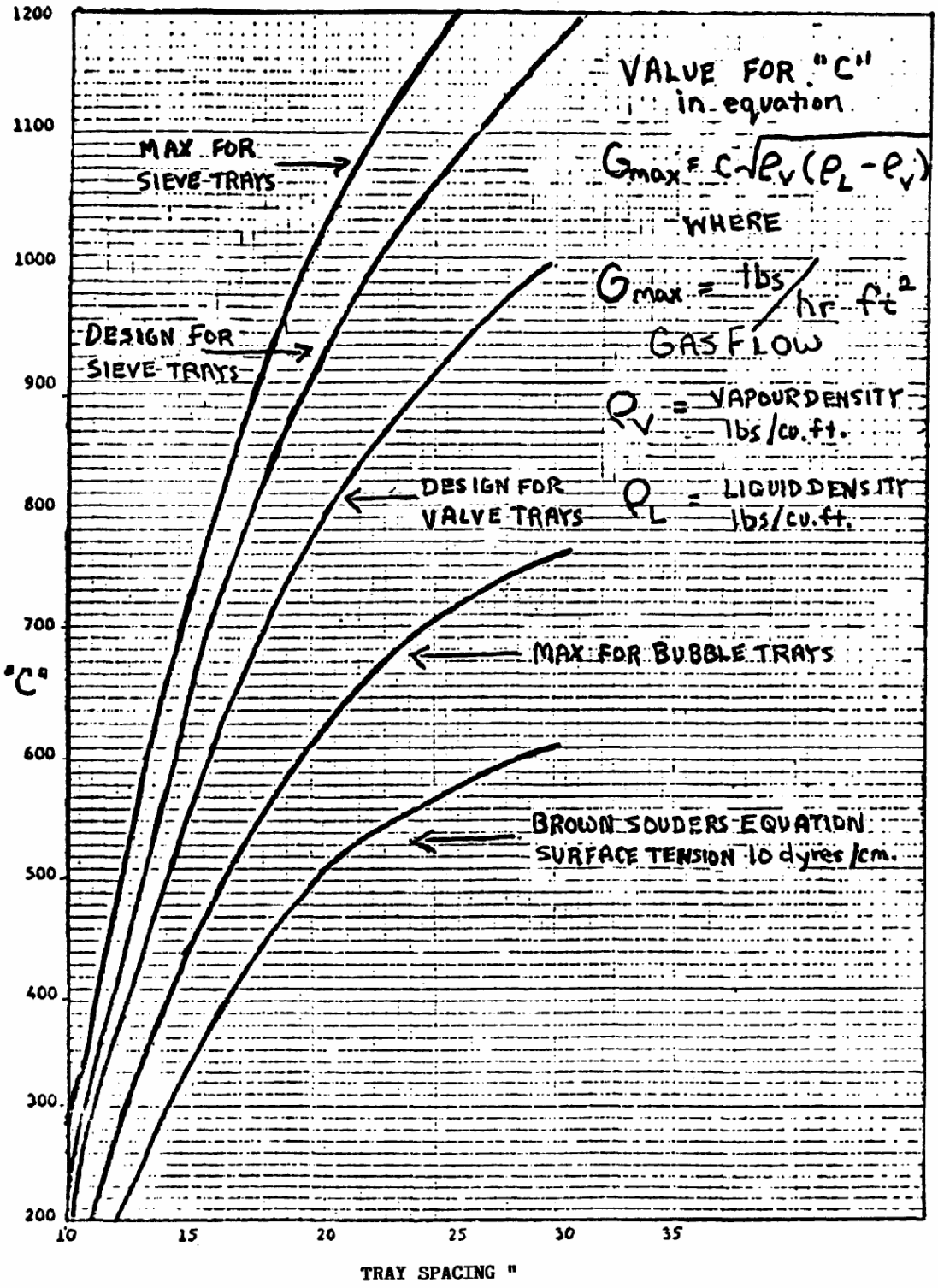


Figure 18.2: Tower Capacity Correlation

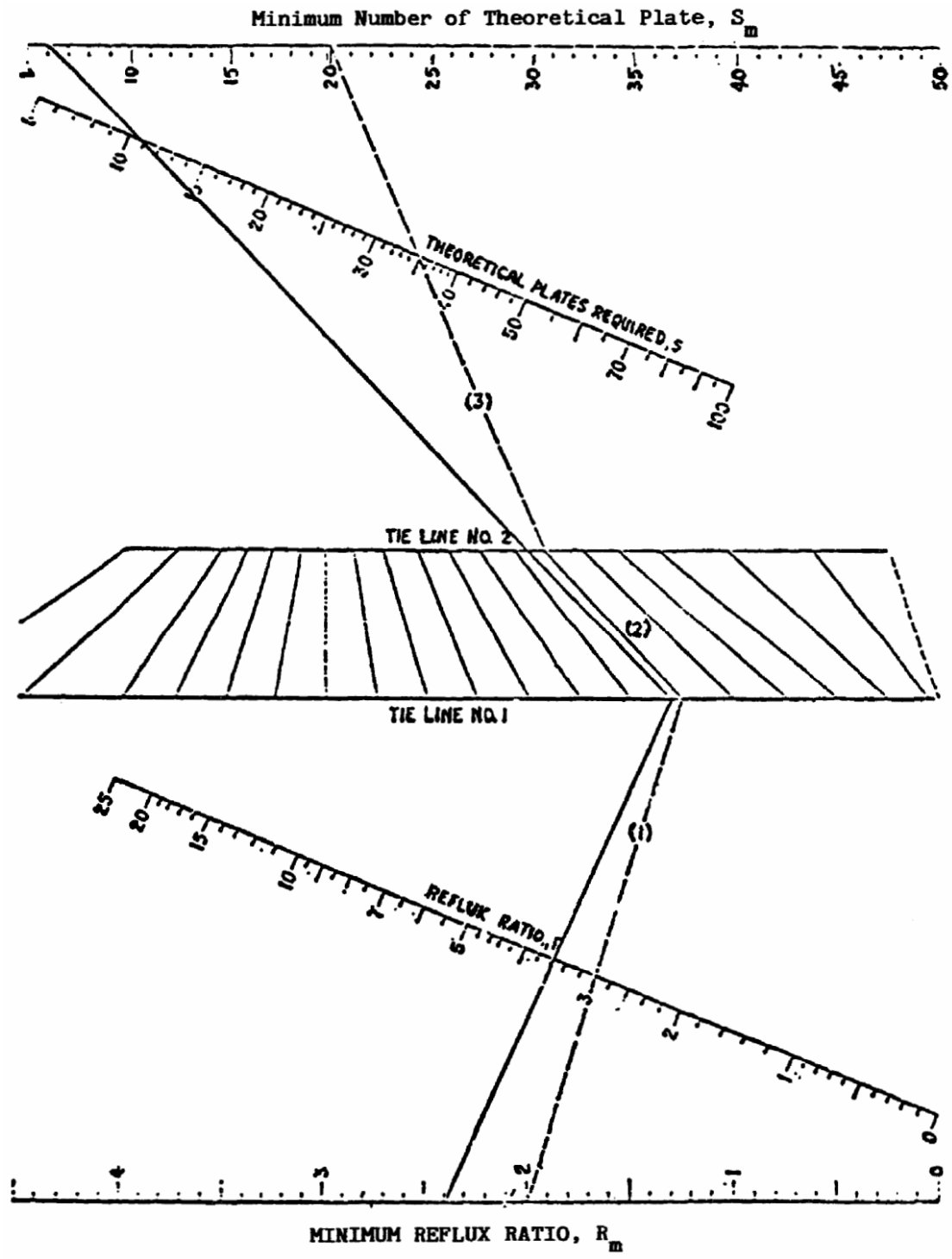


Figure 18.3: Tower Sizing Nomograph

18.2.3 Typical Data on Separations

Tower Service	Actual Number of Trays Required	Actual Reflux	
		Reflux/Product	Reflux/Feed
De-ethanizer	35 - 40	1 - 2	0.5 - 1.0
De-propanizer	30 - 35	4 - 6	1.0 - 2.0
De-butanizer	25 - 30	2 - 3	0.5 - 2.0
Stabilizer	25 - 30	2 - 3	0.5 - 2.0
Butane splitters	70 - 75	10 - 12	4.0 - 5.0
Pentane splitters	79 - 80	10 - 12	4.0 - 5.0

Note: Tray spacing normally 2' or 18" with close clean separation.

18.2.4 Controlling a Tower

To control a tower so that we get a constant product we control:

- the feed rate.
- the operating pressure.
- the reflux rate.
- the heat input.

Note:

- The tower top temperature is a measure of the composition of the overhead product.
- Increasing reflux increases the fractionation and thus the purity of the products.
- Increasing the pressure increases the tower capacity.
- Increasing heat input will increase reflux rate in tower.

Note: Figure 18.4 shows a typical tower control scheme.

Computer controls using composition are now coming into vogue. Primarily based on overhead composition and feed - resets reboil rate.

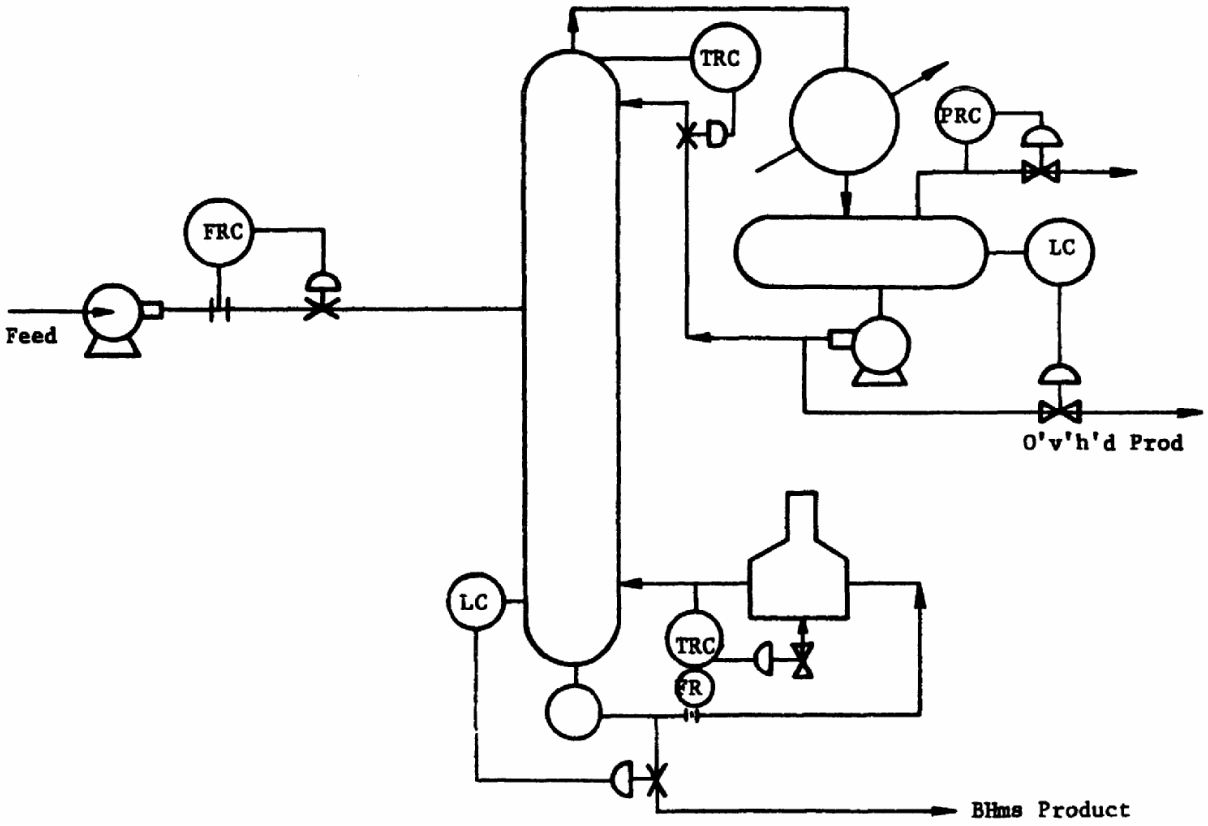


Figure 18.4: Typical Control System

18.2.5 Condensate Stabilization - A Typical Fractionation Process

Figure 18.5 shows the flow sheet for a typical condensate stabilization facility. Briefly, the process is as follows:

> The unstabilized condensate from an inlet separator or from some other hydrocarbon recovery facility is fed to an inlet feed drum, then through heat exchange to a tower where all products lighter than nC_4 are taken overhead and taken off as a gas. The bottoms product which usually contains some nC_4 (up to about 8% in the winter) is cooled and sent to storage. One point to note in this system is the material to use for reboiler types. Carbon steel is often initially specified and then after about 6 months failures occur and alloy tubes are used. Often it is better to start with some sort of alloy tubes, at least 0.5% molybdenum.

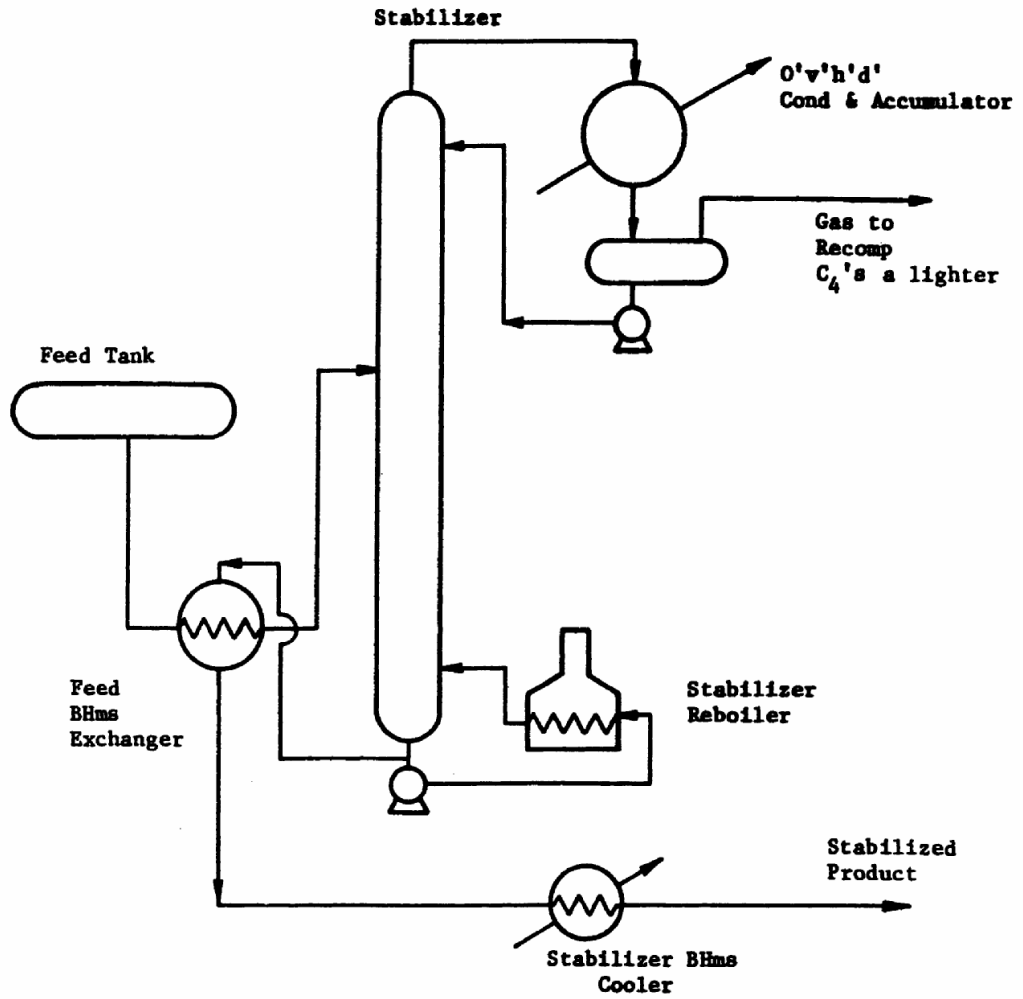


Figure 18.5: Stabilizer Flow Sheet

18.2.6 Example of a Fractionation System

Let us consider the separation of the absorbed material from the dewpoint control plant. The absorbed material will contain C_1, C_2, C_3, iC_4, nC_4, iC_5, C_6^+. The finished products will be:

	Mole %
Fuel gas (C_1 & C_2)	10
Propane	20
Isobutane	10
n butane	15
isopentane	10
n pentane	15
hexane +	20

We can do this directly by eliminating the most volatile components first and continuing on until we have eliminated everything except the hexane. Let us review the process schemes shown in Figure 18.6, Figure 18.7 and Figure 18.8.

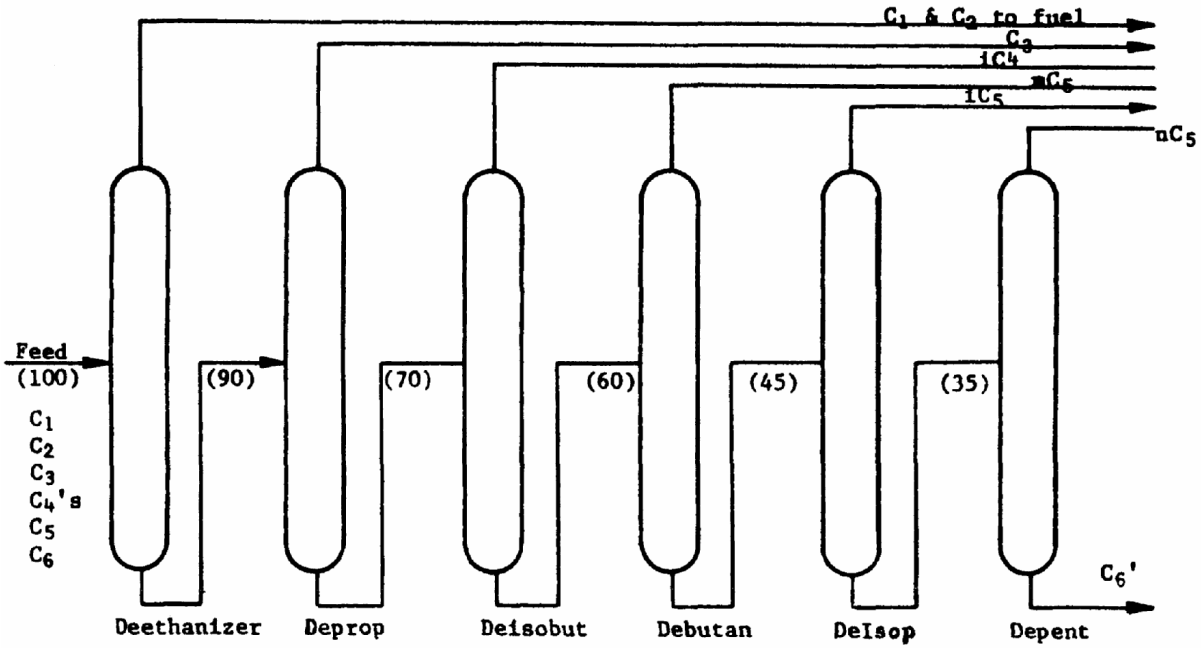


Figure 18.6: Straight Through Processing Scheme – Scheme 1

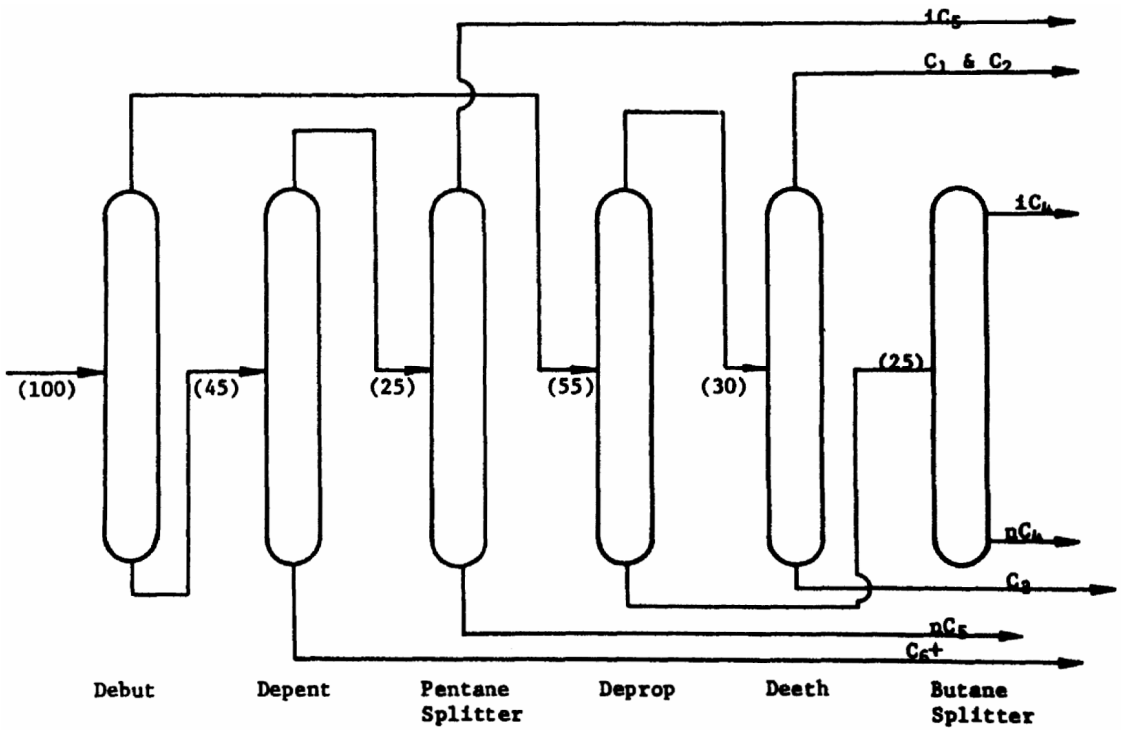


Figure 18.7: Revised Processing Scheme – Scheme 2

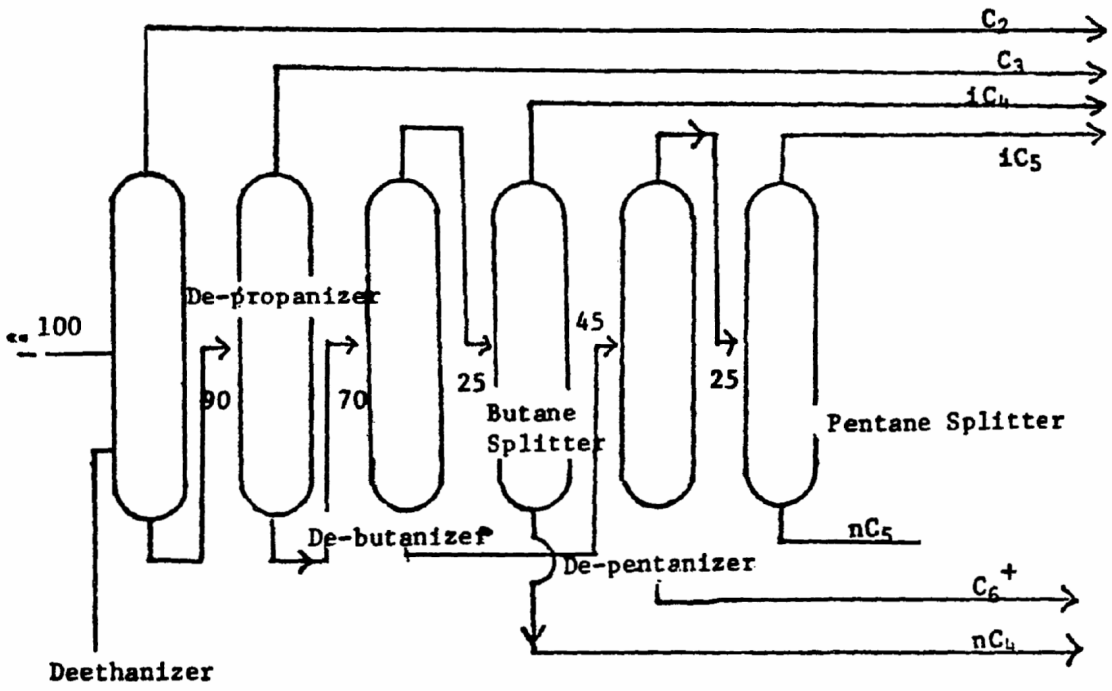


Figure 18.8: Second Revision to Processing Scheme – Scheme 3

a) What is wrong with the process in Figure 18.6?

- Have large volumes when handling close separation, i.e splitting C_4 and C_5 .
- Separations not nearly a 50-50 split.

b) Let us reconsider the process as follows as shown in Figure 18.7:

- Let us split easy components first, for example:

$C_4 + C_5$	$C_3 + C_4$
31 , 80	-40 , +10
BP BP	BP BP

- Then let us fractionate close separations.

c) Lets modify the process scheme in Figure 18.7 slightly so that the close separation are in the tower with two components only, ie. Figure 18.8.

Now let's add up the moles of overhead for each case based on the reflux to feed ratio.

Figure 18.6 Scheme 1	Feed Moles	Overhead Vapour Moles
De-ethanizer	100	100
De-propanizer	90	180
De-isobutanizer	70	350
De-butanizer	60	120
De-isopentanizer	45	225
De-pentanizer	35	70

Total	400	1045
Figure 18.7 Scheme 2	**Feed Moles**	**Overhead Vapour Moles**
De-ethanizer	30	30
De-propanizer	55	110
Butane Splitter	25	125
De-butanizer	100	200
Pentane Splitter	25	125
De-pentanizer	45	90
Total	**260**	**680**
Figure 18.8 Scheme 3	**Feed Moles**	**Overhead Vapour Moles**
De-ethanizer	100	100
De -propanizer	90	180
De-butanizer	70	140
Butane Splitter	25	125
De-pentanizer	45	90
Pentane Splitter	25	125
Total	**355**	**760**

Note: Therefore you should use either Scheme 2 or 3. Since the preliminary overhead vapour moles are relatively close, a detailed balance should now be made between schemes 2 and 3.

18.3 Problems

Example Problem 18.1

It is planned to install a depropanizer to produce product of the split as given in the table below.

	Feed Moles	Ovh'd Moles/100 Mole Feed	Bttms Prod Moles/100 Moles Feed
C_2	4.0	4.0	--
C_3	40.5	39.0	1.5
iC_4	10.5	1.0	9.5
nC_4	16.4	--	16.4
iC_5	10.7	--	10.7
nC_5	8.9	--	8.9
C_6^+	9.0	--	9.0

The minimum reflux ratio for the separation is 1.8 to 1. Please determine how many trays are required in the tower.

Data

- A tower with an actual reflux ratio 1.3 times the minimum costs $500,000. Each tray more than required with the 1.3 times the minimum actual reflux cost $30,000 extra to install including all savings, etc. in the auxiliaries.

- The overhead product to be produced is 25,000 BPD. Extra trays are to be added if they payout in the fuel savings obtained by reducing reflux requirements. The payout time is to be 2 years or less.
- The heat of condensation of a barrel of overhead product is 33,000 BTU's/BBL.
- The cost of fuel is $1.00/MMBTU's delivered to the plant. Assume the fuel is burned in a furnace that is 75% efficient.
- The tower is operating at a pressure of 200 psia. The tower average temperature is 160ºF.
- Assume the tower operates 350 days/year.
- Assume the trays are 80% efficient.

Example Problem 18.2

There has developed a sale of isopentane. The product is not to contain more than 6% mole % of n pentane. An existing tower that has a design operating pressure of 50 psia is available. The auxiliaries are such that the condenser can condense six times the overhead product under normal operation. The minimum reflux needed for the separation is 3.0 to 1.0. The amount of iC_5 in the bottoms product should not be greater than 4 mole %.

Determine the number of actual trays required for the separation. Assume that the fractionator is 75% efficient. The tower feed has the following molal composition.

nC_4	2.0
iC_5	23.2
nC_5	30.8
nC_6	15.6
nC_7	17.4
nC_8	11.0

The average tower operating temperature is 160ºF.

References

1. Chapter 19, 1987 GPSA Engineering Data Book.

19.0 NGL Recovery Processes

- A detailed discussion on the various processes used, their design criteria, application, etc.

19.1 Introduction

LPG's are recovered from most raw gases when the propane and heavier content of a raw gas contains 0.6 U.S. gallons/1000 SCF or if propane content is 1.3 mole % greater for plants up to 200 MSCFD; for plants handling greater volumes of gas C_3^+ contents down 0.5 U.S. gallons/SCF or a propane content of 1.0 mole % should be considered. Also more recently because of the demand for ethane as a petrochemical feed stock and as a miscible fluid for enhanced oil recovery process, a lot of it is now being recovered. If the gas analysis shows the gas is leaner than given above, then one should look very closely at conditions before installing lpg recovery facilities, of course, there are many situations such as large straddle plants, in which the gas is much leaner than given above, but it is economical to recover the lpg's. Each case should be looked at individually. The NGL' are taken out the gas as liquids by:

- Absorption.
- Adsorption.
- Condensation.

19.2 Processes

There are approximately five commercially acceptable lpg recovery processes. Three are used for the so-called Deep Cut Recovery Processes (70% C_3 and 95% C_4 recoveries or greater). However, when ethane recovery is planned generally only the turbo expander process is used. The processes are as given in Table 19.1.

A list of major Canadian plants (as of 1985) is given in Table 19.2 and the process used. We will consider primarily the third and last processes only in Table 19.1. Incidentally all the processes generally have been covered partially in previous chapters.

Table 19.1: Commercial LPG Processing

Process	Application	Examples of Plants in Canada Using Same
Atmospheric Temp. oil absorption.	Low recoveries of C_3 (20 - 30%)	McMahon Plant.
Refrigeration.	Recoveries incidental so that comes out In dewpoint control. Used often in small plants.	Rimbey is now used for small plants particularly for solution gas.
Refrigerated Oil.	Moat common deep cut type plant.	Original Empress and Cochrane plants, Kaybob, E. Calgary, Rimbey.
Adsorption, either charcoal or silica gel, alumina for adsorption.	1 charcoal plant. Several small plants using silica gel.	Caroline in Canada for charcoal. About 8 charcoal adsorption plants. in U.S.
Deep refrigeration with turbo expander.	Most regent progress. Particularly applicable when free pressure available such as on City salee gate but being developed for other services. Good for high ethane recovery. Presently all ethane recovery plants use this process.	Judy Creek III, Strachan, Ricinus, all present Empress plants, New Cochrane plants.

19.2.1 Refrigerated Absorption

19.2.1.1 Description

Figure 19.1 shows the process flow in a typical plant. The gas is first cooled in conventional dewpoint control facilities down to -30 to -35ºF (not so low in rich gas plants). The gas then passes through an oil absorber where the various components are absorbed. The gas leaves the tower and is sent to a presaturator where the oil is saturated with methane and ethane. The gas then is returned to the chilling system for cooling the incoming gas.

The oil that leaves the absorber is called rich oil and contains all the absorbed compounds. The condensed hydrocarbon liquids from the chilling section join this stream. The rich oil is heated and reduced in pressure to about half of the absorbing pressure in the rich oil flash tank. A considerable amount of the absorbed methane is flashed off in this vessel. The oil is then further heated and sent to a de-ethanizer where most of the ethane and the remaining methane are fractionated out of the oil stream. Finally the oil is sent to a HP still where the absorbed products are produced as the overhead product and the bottoms produced is lean oil which is cooled and recirculated to the absorber. The absorbed products are sent to the fractionating facilities.

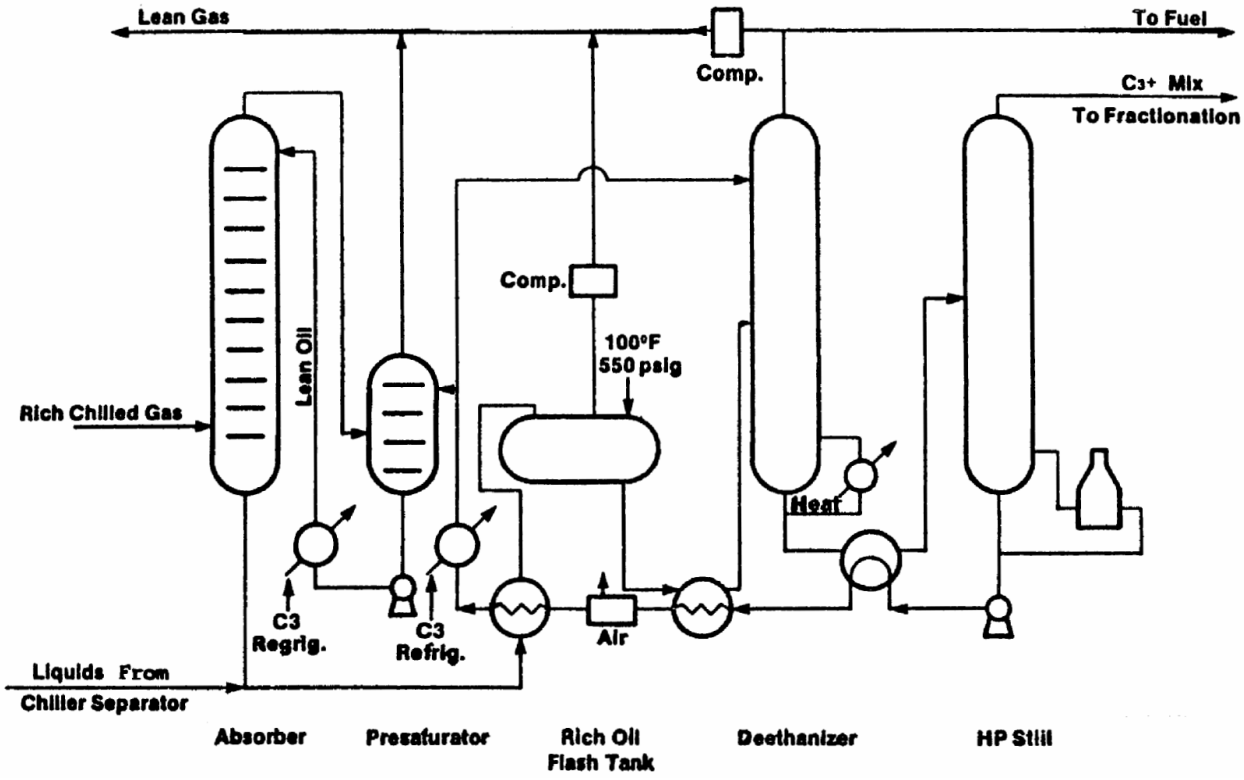


Figure 19.1: Refrigerated Lean Oil System

19.2.1.2 Design Criteria

The primary principle is the design of any absorption unit is how much recovery of material required. The recovery % is set by an economic study of plant investment markets operating costs etc. Generally recoveries of about 80%-85% are used as a design basis.

The oil circulation rate is set by the design recovery and the well known absorption factor " $A = \frac{L}{VK}$ " as discussed in the mass transfer section (Chapter 7). *L* is the moles of liquid circulated. *V* is moles of vapour passing through the tower and "*K*" is the equilibrium constant "K". To find the oil rate required the % absorption of the key component is set (The Y-axis on Fig. 15-51 of the GPA Data book or Figure 19.2 in this section is the % absorbed). Then by travelling across the chart to the number of theoretical trays then to the X-axis (absissa) the absorption factor "A" is found.

Note: Since the gas flow *V* and "*K*" the equilibrium constant is known, then *L* can be calculated and thus the oil rate determined.

Table 19.2: Major Canadian NGL Recovery Plants

Plant Name and/or Location	Operated By	Process Used	Approximate Design Gas Volume m^3x10^6/d	Approximate	
				% C_2 Recovery	% C_3 Recovery
Rimbey	Gulf	Refrig. Oil	11.9	--	95
Stachan	Gulf	Expander	7.7	--	85
Nevis	Gulf	Refrig. Oil	3.5	--	75
Bonnie Glen I	Texaco	Refrig. Oil	3.1	--	75
Bonnie Glen II	Texaco	Expander	4.2	80	90
McMahon Plant, Taylor B.C.	Petro Canada	Atmos Oil	11.1	--	25
Empress	Petro Canada	Expander	56.3	58	95
Kaybob	Petro Canada	Refrigeration	3.1	--	25
Empress	Empress Liquids Joint Venture	Expander	9.8	80	99
Hassar	Canterra	Refrig. Oil	2.4	--	80
Rainbow Lake	Canterra	Refrigeration	2.8	--	50
Harmattan	Mobil	Refrig. Oil	13.9	--	80
Carstairs	Home Oil	Refrig. Oil	9.8	--	80
Ricinus	Amoco	Expander	3.2	--	85
Nipisi	Amoco	Refrigeration	0.8	--	50
Carson Creek	Mobil Oil	Refrig. Oil	2.4	--	80
Ferrier	Amerada	Refrig. Oil	2.8	--	50
0lds	Amerada	Refrigeration	2.4	--	75
Peco	Ocelot	Expander	0.7	80	95
Paddle River	Canadian Occidental	Expander	2.4	--	85
East Calgary	Canadian Occidental	Refrig. Oil	8.9	--	75
Empress I	Dome	Expander	51.0	43	90
Empress II	Dome	Expander	56.6	75	99
Edmonton	Dome	Expander	9.3	90	99
Kaybob I	Dome	Refrig. Oil	6.4	--	85
Kaybob II	Dome	Refrig. Oil	4.8	--	85
Steelman, Saskatchewan	Dome	Expander	0.5	85	99

Plant Name and/or Location	Operated By	Process Used	Approximate Design Gas Volume m^3x10^6/d	Approximate	
				% C_2 Recovery	% C_3 Recovery
Sylvan Lake	Dome	Refrig. Oil	1.8	--	70
Caroline	Dome	Charcoal	1.5	--	90
Caroline	Dome	Refrigeration	1.5	--	50
Garrington	Dome	Expander	0.6	--	90
Goodfare	Dome	Refrigeration	1.9	--	50
West Pembina	Dome	Expander	2.2	--	80
Wembley	Dome	Refrig. Oil	2.8	--	75
Jumping Pound	Shell	Expander	7.2	75	99
Waterton	Shell	Expander	13.3	75	99
Simmonette	Shell	Refrig. Oil	1.0	--	70
Dunvegan	Anderson	Expander	6.0	--	90
Judy Creek I & II	Esso	Refrig. Oil	2.0	--	75
Judy Creek	Esso	Expander	5.0	--	90
Niton	Esso	Refrigeration	1.8	--	60
Leduc	Esso	Refrig. Oil	1.0	--	70
Boundary Lake, B.C.	Esso	Refrig. Oil	0.6	--	75
Quirk Creek	Esso	Refrig. Oil	2.6	--	75
Minnelek - Buck Lake	Esso	Expander	3.0	--	95
Minnehik - Buck Lake	Sulpetro	Refrig. Oil	3.0	--	75
Wapiti	Sulpetro	Expander	2.5	80	99
Elmsworth	Can. Hunter	Expander	11.3	85	99
Kaybob S.	Chevron	Expander	13.8	--	95
Mitsue	Chevron	Refrig. Oil	0.85	--	--
Nevis	Chevron	Refrig. Oil	2.2	--	75
Syivan Lake	Chevron	Refrig. Oil	0.8	--	75
Cochrane	ANG	Expander	31.1	85	99

Example Problem 19.1

Assume you have gas flowing at 100 MMSCFD and that you wish to know the oil rate required for 80% C_3 absorption at -10ºF and 900 psia. The oil has a molecular weight of 120 and an API gravity of 80. The absorber has 6 theoretical trays.

K for C_3 at -10ºF and 900 psia = 0.11

Using Figure 19.2 we find that:

$$"A" = 0.9 = \frac{L}{VK} = \frac{L}{\frac{100 \times 10^6}{379}(0.11)}$$

$$L = 26{,}121\ moles/day$$

$$= 390.6\ US\ gpm$$

This example, of course, is quite a simplified approach.

A more exact method of calculating the lean oil requirement is to use the Edmister method as given in Chapter 7 where an Ae is calculated according to the equation below:

$$Ae = \sqrt{An(A_1 + 1) + 0.25} - 0.5$$

where:

Ae = effective absorption factor.

An = Absorption factor at bottom of tower.

A_1 = Absorption factor at top of tower.

Let us then do the example problem again using this method. We will need further information regarding the system. First we will need the inlet gas composition to the absorber. Assume it is as follows:

C_1	89.6
C_2	7.0
C_3	2.5
iC_4	0.3
nC_4	0.4
C_5^+	0.3

Now to get an 80% C_3 absorption we set up a table to see what amounts of the other components are being absorbed.

Comp.	Analysis	-10ºF K 900	A	% Absorbed	Moles Absorbed per 100m Feed	Comments
C_1	89.6	3.0	0.033	3.7	2.95	
C_2	7.0	0.42	0.235	23.5	1.65	
C_3	2.5	0.11	0.900	80	2.0	Read 'A' required for 80% from chart.
iC_4	0.3	0.044	2.25	100	0.3	
nC_4	0.4	0.031	3.19	100	0.4	
C_5^+	0.2	0.003	33.0	100	0.2	Use C_6 Values.
Total					**7.50**	

Find L/V from $A = \frac{L}{Vk}$ for C_3 = 0.9 (.11) = .099 and use this to calculate the rest of the components absorbed.

Terminal Conditions

$$Inlet\,(Rich)\,Gas = \frac{100 \times 10^6}{379} = 263{,}852\ moles$$

$$Outlet\,(Lean)\,Gas = \frac{100 \times 10^6}{379}(100 - 7.5) = 244{,}406\ moles$$

$$Lean\,Oil = X = X\ moles$$

$$Rich\,Gas = X + 0.075\frac{(100\times 10^6)}{(379)} = X + 19{,}789\ moles$$

Solve Edmister equation putting in A1 & An & Ae with X in same:

$$0.9 = \left[\frac{X + 19{,}789}{(263{,}852)(0.11)}\right]\left[\left[\frac{X}{(244{,}406)(0.11)}\right]+1\right] + 0.25 - 0.5$$

$$X = 15{,}600\ moles/day$$

Thus would require 60% of the volume of lean oil for the required absorption as compard to previous calculation.

The number of trays in the absorber is usually 25 to 35. The efficiency of absorbers varies with the "key" component (ie. the principal one that is being absorbed). It is generally between 25% and 35%.

The oil circulation is of course also a function of the absorption temperature. The lower the temperature, the higher the absorption and thus less needs to be circulated. The Table below indicates the molecular weight of the oil used at various temperatures.

Table 19.3: Absorption Oil Molecular Weight

Temp °F	Mol. Weight
+80	180-200
0	140-160
-30	100-120

The lower the molecular weight of the oil, the more that is lost overhead from the absorber. This can sometimes be prevented by using a sponge oil section on top of the tower with a heavy oil in it to recover only the light regular absorption oil.

As there is a large heat of reaction and as there is also a fair bit of light component absorption often a presaturator is put in the oil circuit. This is a short tower with 4 to 6 trays. In C_3 absorption plants, a fair bit of the methane and ethane are absorbed here and the heat of absorption removed by further chilling.

One of the important design variables in the system is of course the "K" value chosen. The selection of the appropriate "K" value for a system is discussed in some detail in the Vapour-Liquid Equilibrium Chapter of this course (Chapter 4). The NGPA has been running tests of "K" values of oils for some time. This work was reported by Hwang, Erbar & Robinson.

The deethanizer and the HP Still are sized by normal fractionation methods. The deethanizer has usually 35-45 trays. The Still usually has about 25-30 trays.

The other main design consideration is a good heat balance. This is essential as high temperature is required to regenerate the oil and low temperature is required for absorption. Close approaches are used to minimize refrigeration, air cooling and hot oil heating requirements.

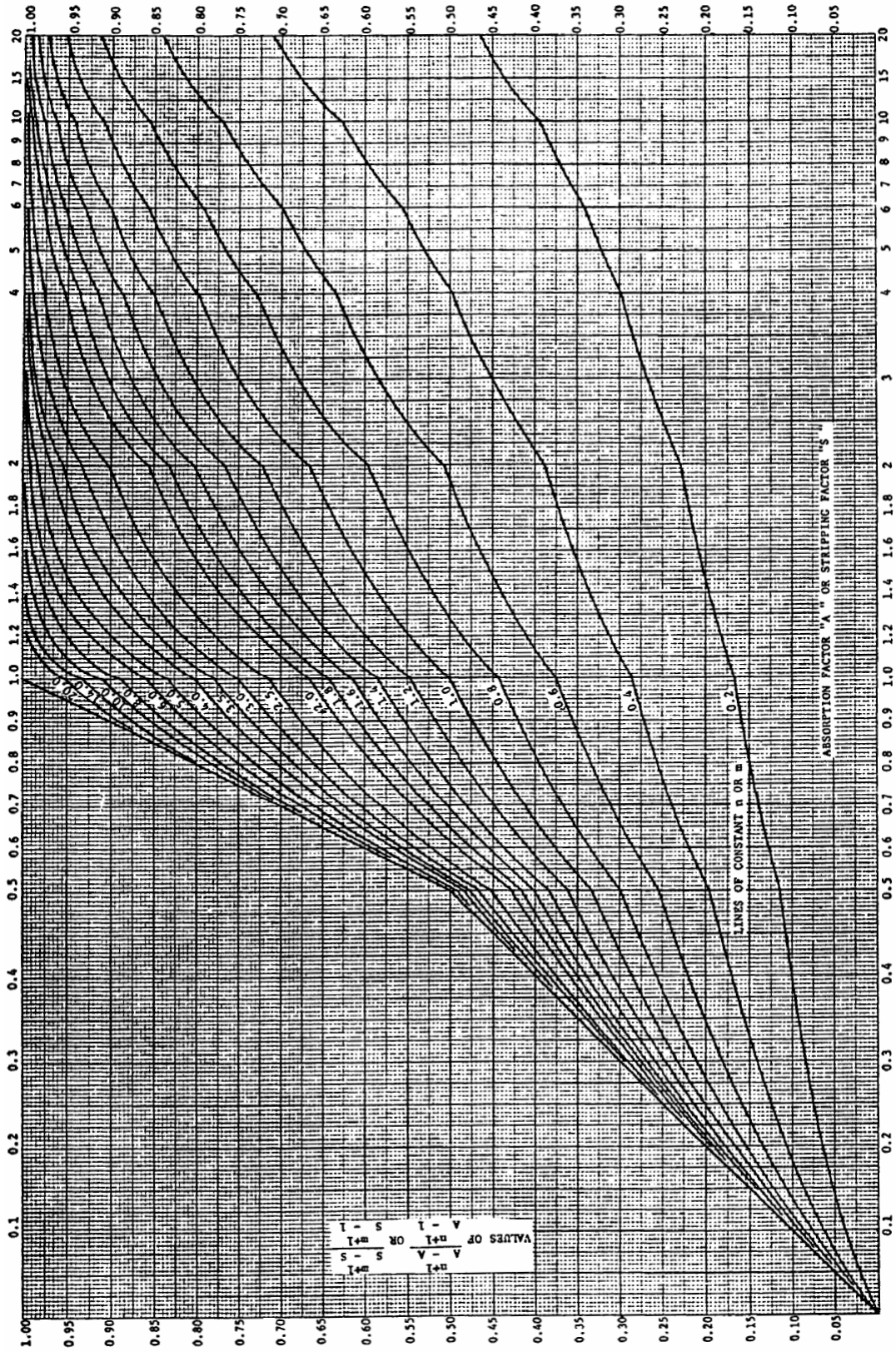


Figure 19.2: % Absorbed Vs Absorption Factor

19.2.1.3 Operations

Absorption plants generally run quite well. Corrosion is essentially minimum. The major problems are with the actual mechanical equipment such as pumps, compressors etc.

The lean oil which is usually made up from the gas must be kept dry and the heavy ends must be fractionated out of it.

Foaming is normally not the problem that it is in sweetening units. However, aromatics or cracked materials in the lean oil sometimes cause it to occur.

19.2.2 Adsorption

For deep cut plants, only carbon adsorption plants have been used to any extent. These plants have had some design problems, but properly designed, the plants seem to work reasonably satisfactorily. The particular problem is the wide temperature variations experienced between adsorption and regeneration. Two processes using charcoal as the adsorbent are commercially used. These are:

- Kilgore and Nasser Process.
- The Continental Oil Process (This is discussed in the article by C.A. Barrere).

19.2.2.1 Process Description

The Continental Process will be described. Figure 19.3 shows the process. This process is slightly different in that closed recycle streams are used for regeneration and cooling. The process is quite simple. One bed is on the absorbing cycle, one bed is regenerated and one bed is cooled. The thermal or volume compensators are used to balance the pressure in the regeneration and cooling cycles.

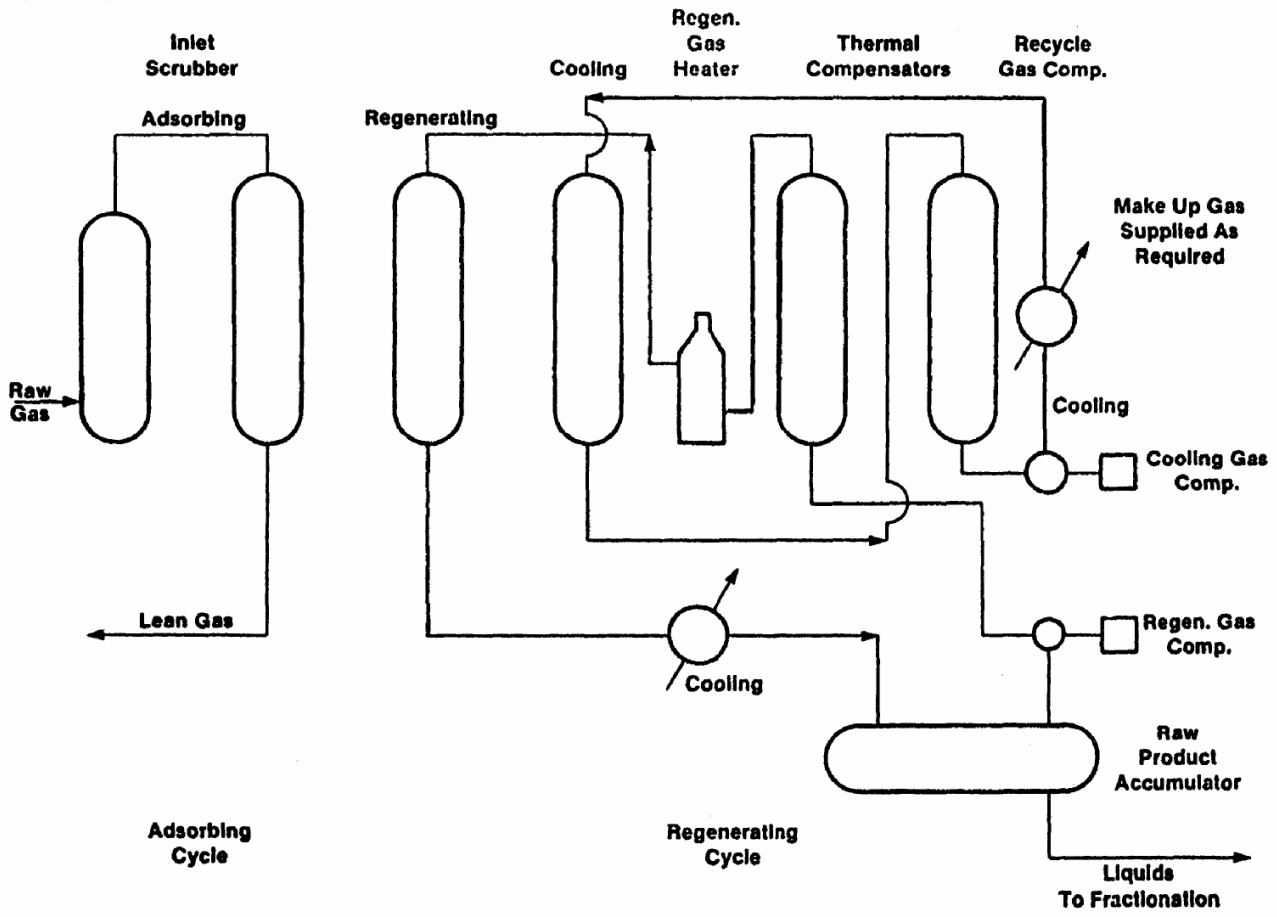


Figure 19.3: Adsorption Process

19.2.2.2 Design Data

Some discussion has taken place on the design requirements in the section on Dewpoint Control, on such items as cycle time. For determining the bed length, it is necessary to refer to experimental data. Two references for doing this should be consulted:

- Humphries, C.L., for Silica gel type adsorbents.
- Barrere, C.A. , for carbon adsorbent.

The particular problem is that the amount of adsorbent required is a function of the gas composition and thus experimental data is required.

Generally the lbs of adsorbate per 100 lbs of adsorbing material is a function of the inlet composition, the pressure, the temperature and the velocity of the material through the bed. The value is different for each component. Figure 19.4 gives some data for C_5 and C_6 adsorption on silica gel at 800 psig and 90ºF.

The method of sizing the adsorption bed is done by adding up the total adsorbent required for each component. The diameter and height is determined by the fact that the normal superficial velocity through the beds is set at about 45 ft per minute. It should be noted that composition of the gas affects the adsorption capability for each inlet component.

For regeneration, it is necessary to take the total adsorber up to the regenerating temperature. This normally takes at least 20 minutes. This then sets the minimum cycle time. A slightly longer time can of course be set.

The regeneration temperature is about 700ºF.

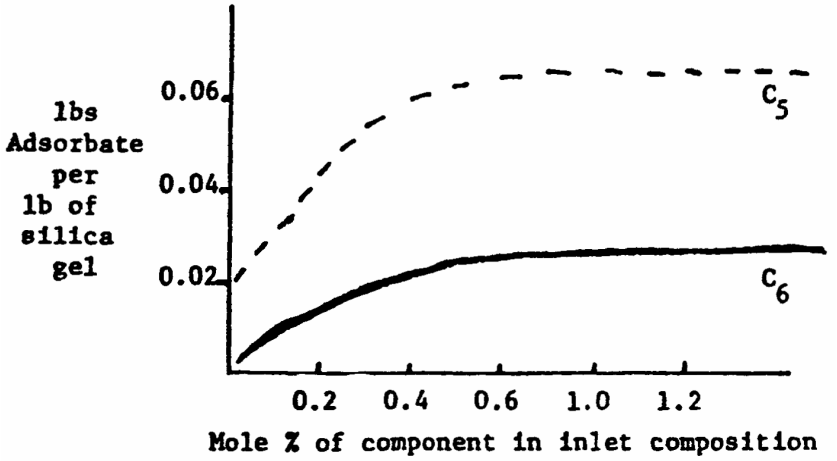


Figure 19.4: Adsorption of C_5 and C_6

19.2.2.3 Operation

Approximately 8 carbon adsorption plants are in service to recover propane and heavier (Two of these actually recover up to 50% ethane). Three of these plants are working with the Kilgore patent. In one of these plants, severe mechanical problems developed in high stresses due to changes in temperature when switching from adsorbing to regeneration and return. Also, the carbon beds were lost initially. The Conoco process has not experienced mechanical problems but has had problems with getting design recoveries. Heavy ends like compressor lube oils, etc., can cause problems with the adsorbing beds.

19.2.3 Low Temperature Refrigeration (Cryogenics or Expander Plant)

19.2.3.1 Process Description

Figure 19.5 shows a typical process flow for an expander type plant. The raw gas enters into an inlet separator where any liquids are separated (ie. lube oil, etc.). The gas is dried to about -150ºF dewpoint, (about 10ºF below the lowest expected temperature). The gas is then cooled in gas-to-gas exchangers. Condensed liquids are removed and then the gas is sent to an expander. This is a turbine that reduces the pressure to about half the entrance pressure and cools the gas by taking work out of it (down to -130ºF or lower). The liquid hydrocarbons are again separated and then fed to a demethanizer in which the methane is removed, then to the de-ethanizer for production of specification ethane and propane$^+$. The lean gas is used, along with the other liquids, to cool the incoming gas. It is then recompressed to the required outlet pressure using the energy available in the pressure drop and considerable additional energy from outside to get it back to its original pressure.

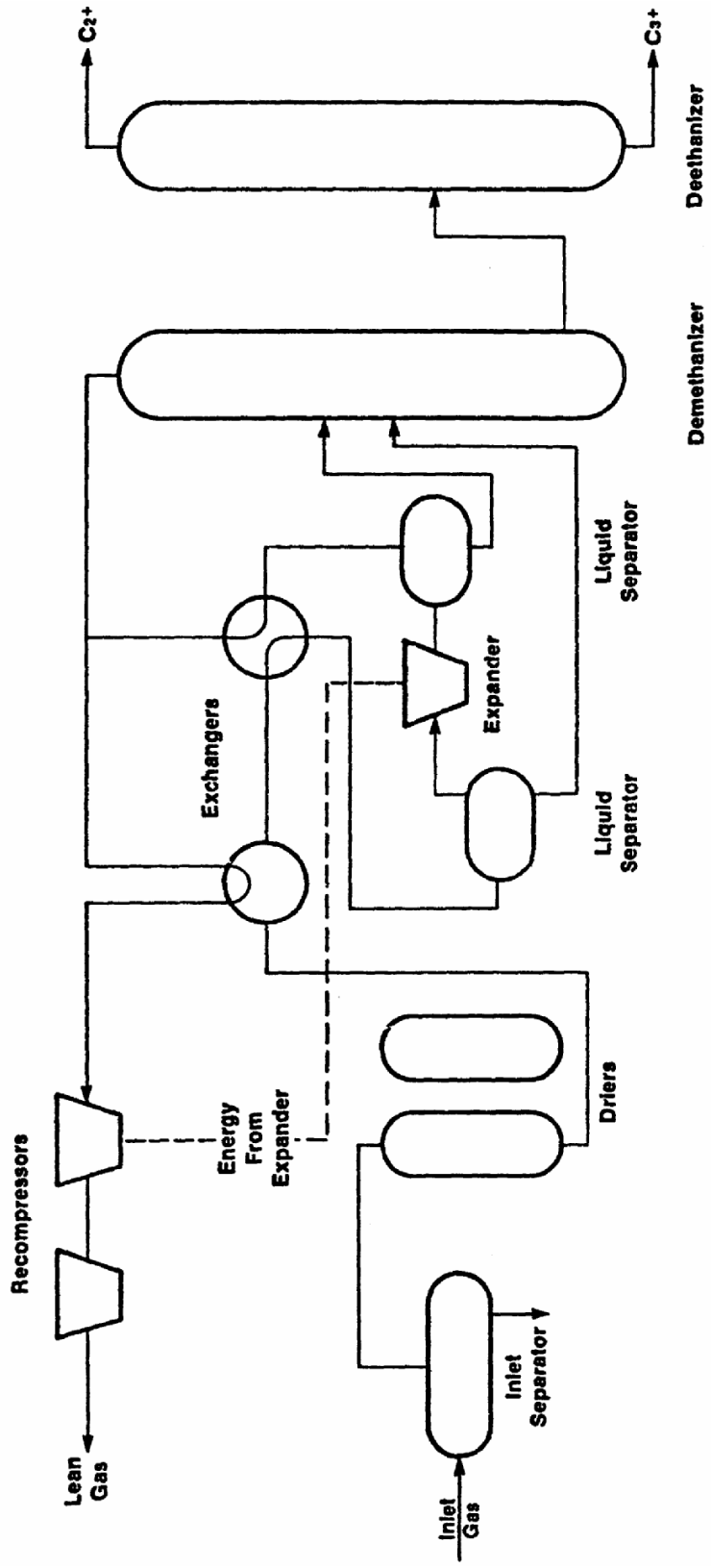


Figure 19.5: Expander Process

19.2.3.2 Process Design

The main feature of the expander is to chill the gas as low a temperature as possible so as to condense as much of the heavy components as possible. Accurate equilibrium data "K's" is very important, a slight change in temperature changes the amount of condensation of methane particularly. The extra condensing of methane often gives problems in the demethanizer. The important feature is that the system be designed so that the outlet of the expander is well within two phase zone. Chapter 10 of these notes and Pages 5-11, 5-12, and 5-13 of the GPA data book explain the calculation of the expander outlet temperature. Figure 19.6 shows the pressure temperature relationships in the process. The reduction in the phase envelope is brought about by condensation before the expander.

The original economics of product recovery were based on the horsepower required for the process. The lower the temperature the greater the liquid recovery but this of course requires a lower pressure and thus greater horsepower for recompression of this is required for redelivering the lean gas. Increased recovery by refluxing the demethanizer or cold oil absorption has now been used to material increase the ethane recovery without a material increase power (see Milne, and Wilkson and Hudson, also Ortloff Patent). The processes for increased ethane are discussed further on in this chapter.

Generally the pressure reduction is about 1/2 to 1/3 of inlet pressure and the enthalpy drop across the expander is around 50 BTU's/lb (116 Kj/Kg) of gas flowing.

The metallurgy has to be carefully selected as temperatures in the -160ºF (-107ºC) or lower are often experienced. Nickel steel, Stainless Steel or Aluminum are used at these lower temperatures.

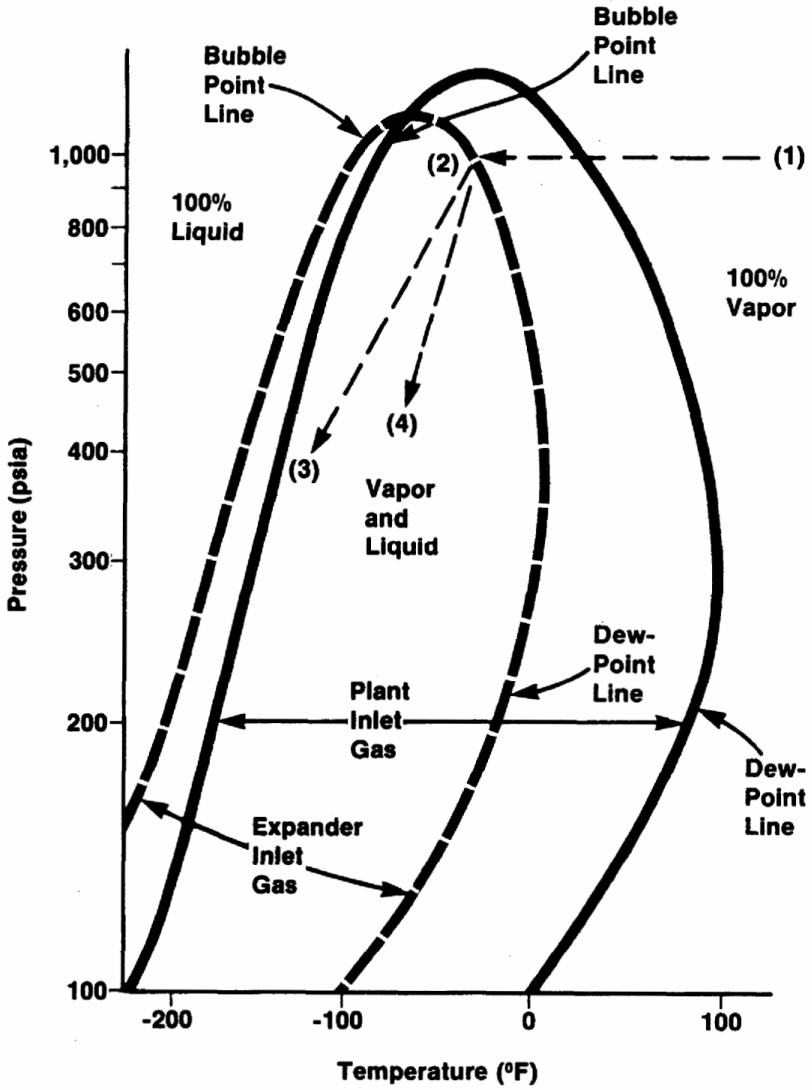


Figure 19.6: Pressure Temperature Diagram Expander Plant

19.2.3.3 Operations

All reports of operations are generally good, including the operation of the expander which runs up to 15,000 rpm. However, some problems with hydrates and seals leaking have been reported. Hydrates forming on rotor blades have caused a problem. Good dehydration is essential. Low CO_2 of inlet gas is desirable as solid CO_2 can give problems (see Figure 5-17 and write-up in GPSA Manual). An expander can handle up to 25% liquids formed in it. Morgan Farun has compared the turbo expander process for recovering lpg's with other processes. Cochrane has described the Empress Expander plant quite well. Also Parker gives a good description of the operation of an expander plant. Morgan, Gulsby and Han et a describes some expander plants. Table 19.2 lists the plants presently using the expander process in Canada.

Please refer to “Chapter 10.0 Compression and Refrigeration” for enthalpy and entropy calculations.

To increase the recovery of ethane processes have been developed to reabsorb the ethane and any propane that was leaving the basic system in the gas being returned to the pipeline (In Figure

19.5 this would be the gas streams leaving the LP Separator after the expander and the demethanizer).

Two ways are used to absorb the ethane. The first method is to use a cold liquid stream made up of feed gas. This method is shown in Figure 19.7. The second method provides the absorbing fluid by chilling and condensing residue gas being returned to the pipeline. The flow sheet for this method is shown in Figure 19.8. Please note both of these flow sheets depict the cases when the equipment additions are being made to an existing plant. If a plant is being designed from the beginning often the cold reflux (absorption fluid) is incorporated in the top of the demethanizer.

Sometimes both systems of supplying a cold absorption oil are used (ie. feed liquids or residue gas liquids).

The recovery of ethane and propane can be calculated like a regular absorber that was discussed earlier in this chapter.

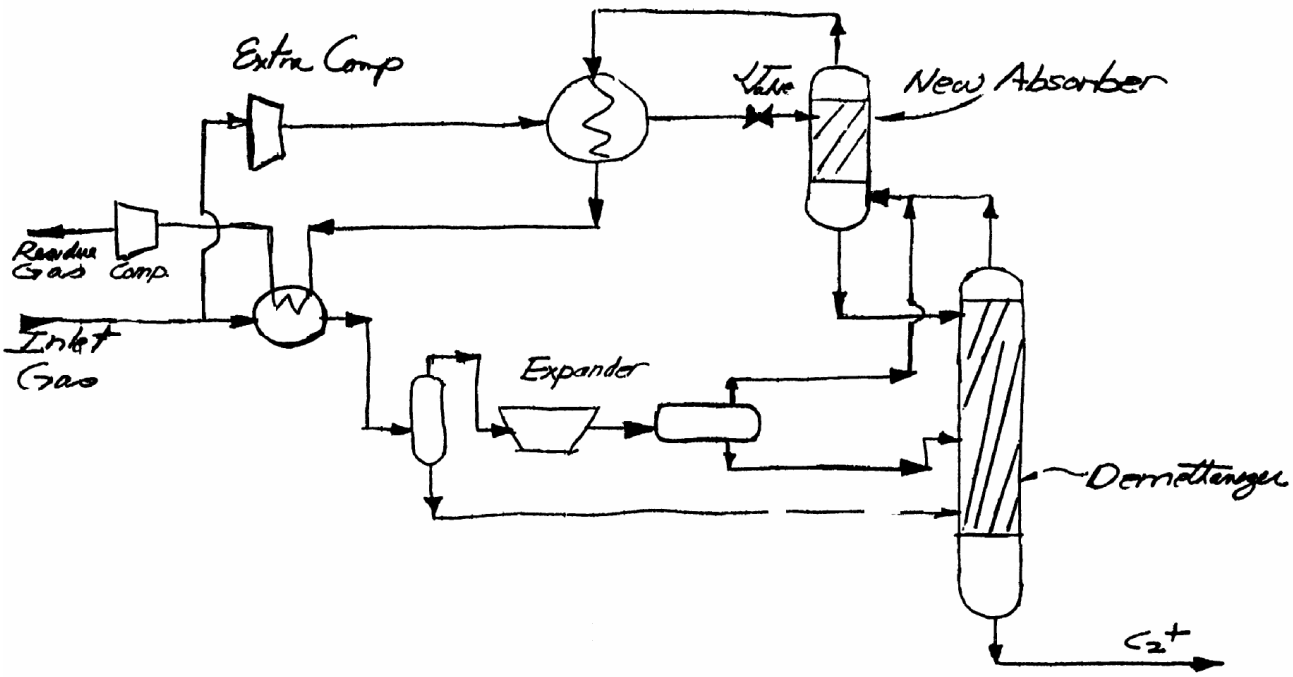


Figure 19.7: Feed Condensation Scheme

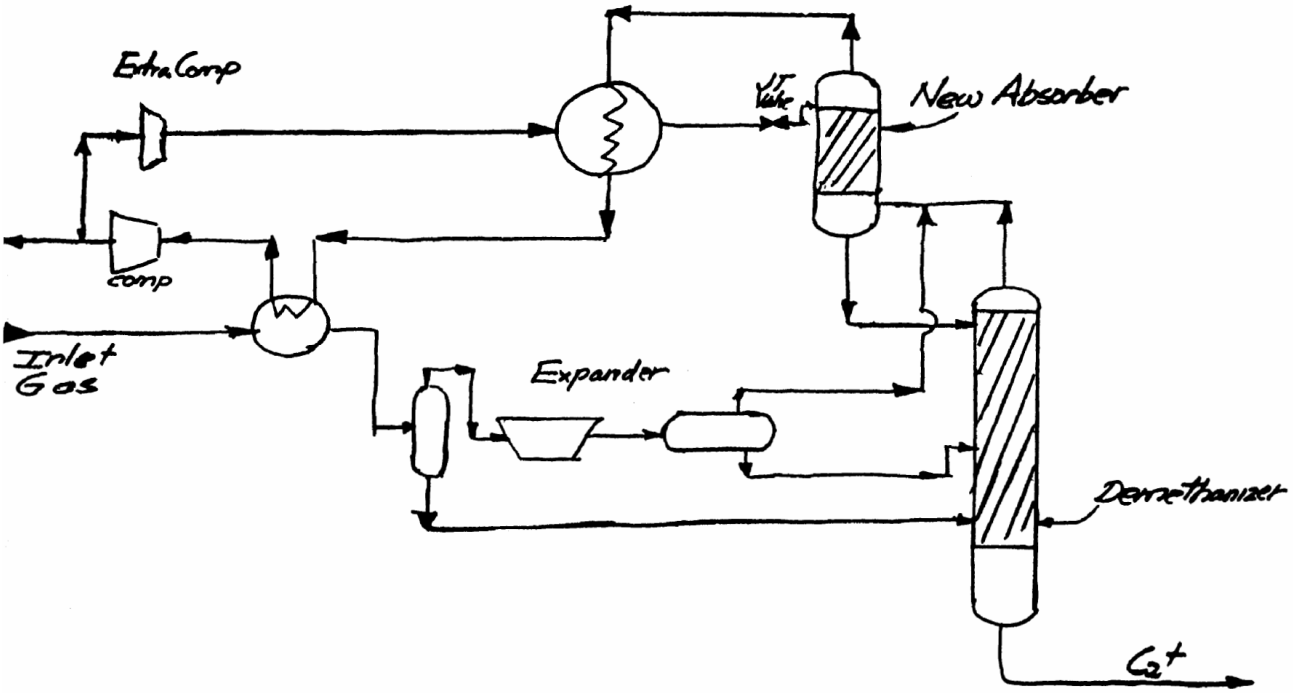


Figure 19.8: Residue Gas Condensation Scheme

19.3 Problems

Example Problem 19.2

A gas plant is handling 1 BSCFD of gas of the composition given below:

Comp.	Mole %
N_2	0.3
CO_2	0.2
C_1	89.9
C_2	6.5
C_3	2.1
iC_4	0.3
nC_4	0.5
C_5	0.2

The oil absorption process is used. Normally the absorption takes place at an average of -35ºF. The recovery at this temperature is 85% of the C_3 and essentially all the iC_4 and heavier. A mechanical failure reduces the availability of refrigeration. Two choices of operation are possible.

1. Run only 600 MMSCFD through the plant and keep the same temperature of absorption, ie. -35ºF and the same oil rate (same as original) or
2. Keep the total gas flow through the plant (ie. 1 BSCFD) and reduce the temperature of absorption to 0ºF with same oil rate (same as original).

Data

- The operating pressure is 500 psia.
- Assume each case operating costs are the same.
- Assume the values for the products are:

  C_3 = \$18.00/BBL

  iC_4 = \$25.50/BBL

  nC_4 = \$23.00/BBL

  C_5 = \$36.00/BBL
- Assume the absorber has 4 theoretical trays.

Determine which method to use.

Example Problem 19.3

You are to determine the pressure that the expander outlet must be set so that the ethane recovery from the gas that is fed to it is at least 60%. The gas fed to the expander is at -65ºC and 6000 kPa and has a composition given below: (It is all gas at this condition).

Composition	Mole %
C_1	91.0
C_2	6.0
C_3	2.5
iC_4	0.2
nC_4	0.3
Vapour Entropy kJ/Kilo-mole ºC	140.6

Assume all the ethane that is in the liquid phase in the LP separation after the expander is recovered.

Here is some thermodynamic equilibrium data for your use:

Pressure 3000 kPa					
	K-Values				
Temp. -ºC	**80**	**85**	**90**	**95**	**100**
C_1	1.53	1.34	1.17	--	--
C_2	0.132	0.118	0.110	--	--
C_3	0.026	0.026	0.028	--	--
iC_4	0.008	0.010	0.016	--	--
nC_4	0.005	0.007	0.013	--	--
Vapour Mole %	82.5	73.0	46.3	0	0
Liquid Mole %	17.5	27.0	53.7	100	100
Vapour Entropy kJ/km/ºC	142.0	140.0	138.0	--	--
Liquid Entropy kJ/km/ºC	134.0	131.0	127.5	--	--
Pressure 2500 kPa					
	K-Values				
Temp. -ºC	**80**	**85**	**90**	**95**	**100**
C_1	1.85	1.72	1.48	1.35	--

C_2	0.132	0.115	0.101	0.09	--
C_3	0.022	0.020	0.018	0.016	--
iC_4	0.006	0.005	0.004	0.004	--
nC_4	0.004	0.004	0.003	0.003	--
Vapour Mole %	87.0	82.5	65.8	40.0	--
Liquid Mole %	13.0	17.5	34.2	60.0	--
Vapour Entropy kJ/km/ºC	146.0	144.1	142.0	140.5	100
Liquid Entropy kJ/km/ºC	130.0	129.2	128.6	127.1	--
Pressure 2000 kPa					
	K-Values				
Temp. -ºC	**80**	**85**	**90**	**95**	**100**
C_1	2.39	2.07	1.79	1.53	1.28
C_2	0.133	0.112	0.093	0.081	0.070
C_3	0.018	0.015	0.013	0.012	0.011
iC_4	0.004	0.004	0.004	0.005	0.008
nC_4	0.002	0.002	0.002	0.004	0.007
Vapour Mole %	91.7	89.1	85.3	79.0	64.3
Liquid Mole %	8.3	10.9	14.7	21.0	35.7
Vapour Entropy kJ/km/ºC	152.0	148.4	145.0	142.9	140.5
Liquid Entropy kJ/km/ºC	140.0	137.3	134.5	132.1	130.0

Note: K-values have been rounded off.

Note: km means kilo-mole.

Assumptions

To take care of the expanders efficiency, allow a 1.5 kJ/km entropy increase.

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Reading List

1. Transcripts of (GPSA Seminar - Sept. 19, 1985), "NGL Recovery Plant Update Design and Operating Experience". Summary published in Oil & Gas Journal and Energy Processing.
2. The 1987 GPSA Data English Version Chapter 19.

20.0 Sulphur Recovery

- A review of the various processes used to recover sulphur
- How they are designed and operated.
- *This chapter was revised using extensive notes developed by H. Paskell of Western Research.*

20.1 Processes

There is one process that is mostly used. This is the so-called Modified Claus Process (developed by Claus in 1890 and subsequently modified and commercialized by I.G. Farben Industries in 1937). This process consists of burning 1/3 of the H_2S to SO_2 in a reaction furnace. The following equations represent the reaction that takes place:

Reaction 20-1

$$H_2S + 3/2\,O_2 \rightarrow H_2O + SO_2 - H$$
$$H = 124 - 133\ Kcals$$

The reaction products are cooled and the rest of the H_2S is reacted with the SO_2 over a catalyst according to the following reaction:

Reaction 20-2

$$2H_2S + SO_2 \rightarrow 2H_2O + \frac{3}{\ell}S_\ell - H$$
$$H = 21 - 35\ Kcals$$

where:

ℓ = the species of S formed.

or overall:

Reaction 20-3

$$3H_2S + 3/2\,O_2 \rightarrow 3H_2O + \frac{3}{\ell}S - H$$

Figure 20.1 shows a typical flow sheet for the process. Basically the claus unit can be divided into 6 primary operations:

1. Combustion.
2. Gas Cooling.
3. Reaction of H_2S and SO_2.
4. Condensation of elemental Sulphur.
5. Reheating of Reactor feeds.
6. Incineration of the tail gas.

We will discuss this process in more detail later.

To handle various % H_2Sin the acid gas various modifications of the Claus process are used. These are discussed later.

For small plants in which the recovery of commercial sulphur is not as important as the actual removal of sulphur from the gas the so called emission plants are used. These processes depend on converting H_2S directly to elemental sulphur and the sending of the elemental sulphur either to a landfill or a re-melter. The most common of these processes is the Stretford and the "Lo-Cat". These have been discussed briefly in Chapter 15 and will be discussed later in this Chapter.

Table 20.1 gives a list of the major sulphur recovery facilities in Western Canada in 1986.

Table 20.1: Major Sulphur Processing Plants - 1986

Area	Operator	Sulphur Capacity tonnes/day
Olds	Amerada	389
Bigstone	Amoco	385
Crossfield	Amoco	1797
Crossfield	Canadian Occidental	1696
Mazeppa	Canadian Occidental	577
Rainbow	Canterra	139
Ram River	Canterra	4572
Brazeau River	Canterra	42
Okotoks	Canterra	431
Windfall	Canterra	1199
Kaybob S. No. 3	Chevron	3557
Sinclair (Hythe)	Chieftain	256
Sturgeon Lake	Dome	98
W. Pembina	Dome	347
Zama	Dome	74
Brazeau River	Dome	110
Edson	Dome	288
Kaybob S. No. 1	Dome	1086
Kaybob S. No. 2	Dome	1086
Lone Pine	Dome	283
Quirk Creek	Esso	299
Strachan	Gulf	953
Hanlan	Gulf	1092
Homeglen Rimbey	Gulf	128
Nevis	Gulf	299
Carstairs	Home	65
Wimborne	Mobil	182
Harmattan	Mobil	515
Lone Pine	Mobil	157
Wildcat Hill s	Petro-Canada	177
Brazeau River	Petro-Canada	78

Area	Operator	Sulphur Capacity tonnes/day
Gold Creek	Petro-Canada	43
Innisfail	Shell	163
Waterton	Shell	3107
Jumping Pound	Shell	597
Burnt River	Shell	489
Rosevear	Shell	171
Simonette	Shell	95
Ninnehik - Buck Lake	Sulpetro	45
Rosevear	Suncor	110
Coleman	Westcoast	389
Fort Nelson, B.C.	Westcoast	110
Pine River, B.C.	Westcoast	1055
Taylor, B.C.	Westcoast	460

20.2 MODIFIED CLAUS PROCESS

Let us discover this process in some detail. The flow sheet given in Figure 20.1 shows the straight through type process in reasonable detail. This process consists of burning the H_2S to $S0_2$ in the reaction furnace, cooling the reacting gases to condense out the sulphur formed, then reheating the reactants SO_2 and H_2S to approximately 450ºF and passing them through a catalyst chamber so that the Reaction 20-2 given previously occurs. Then again cooling the gases to condense the sulphur and again repeating the reheating, catalytic reaction and cooling sequence.

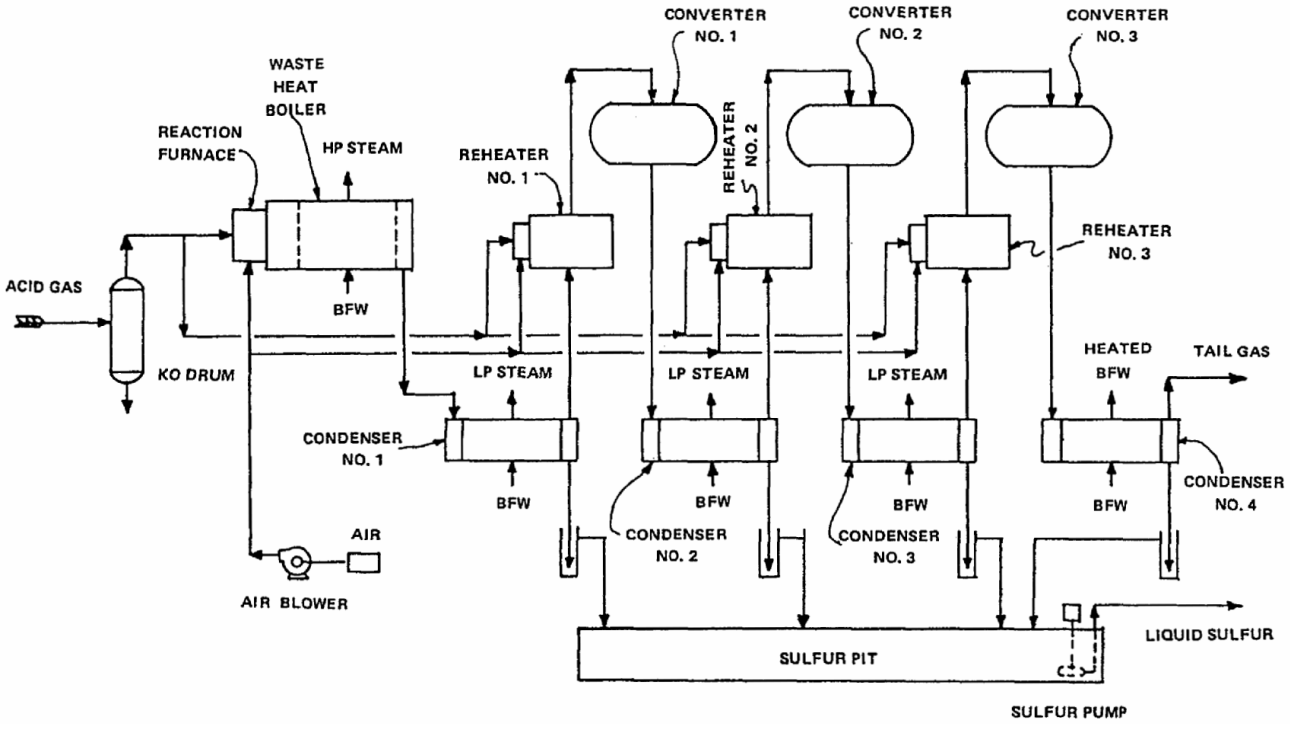


Figure 20.1: Claus Sulphur Plant

The method of reheating is one of the major differences in straight through process.

The methods commonly used are:

a) Reheating by means of auxiliary burners.

b) Reheating indirectly by means of outside gases and/or reaction gases.

c) Reheating by means of hot gas by-passes.

c) is generally the one that gives the best recovery but is the most expensive. a) gives almost as good recoveries and b) also gives almost as good recoveries and is the cheapest. The processes are shown on Figure 20.2 and Figure 20.3.

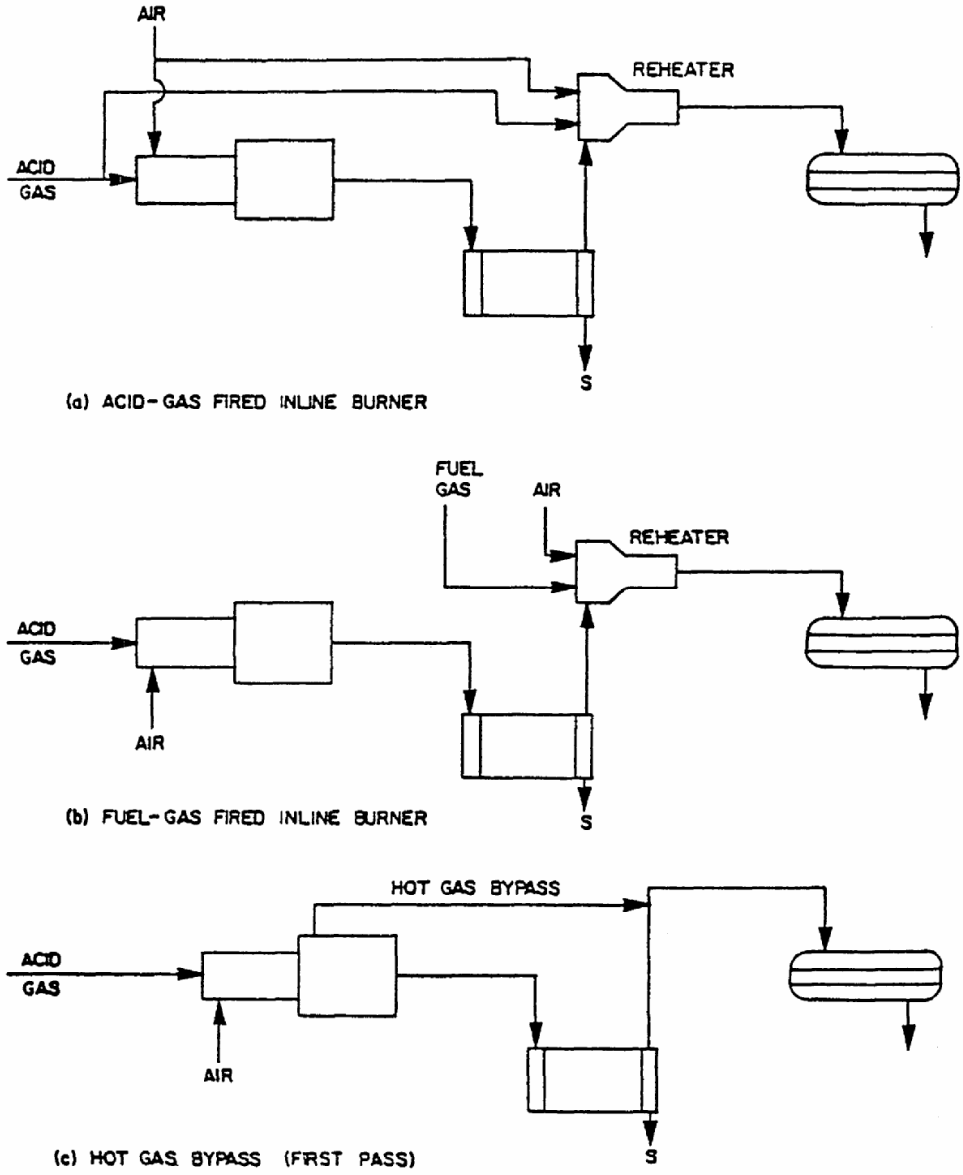


Figure 20.2: Direct Heat Methods

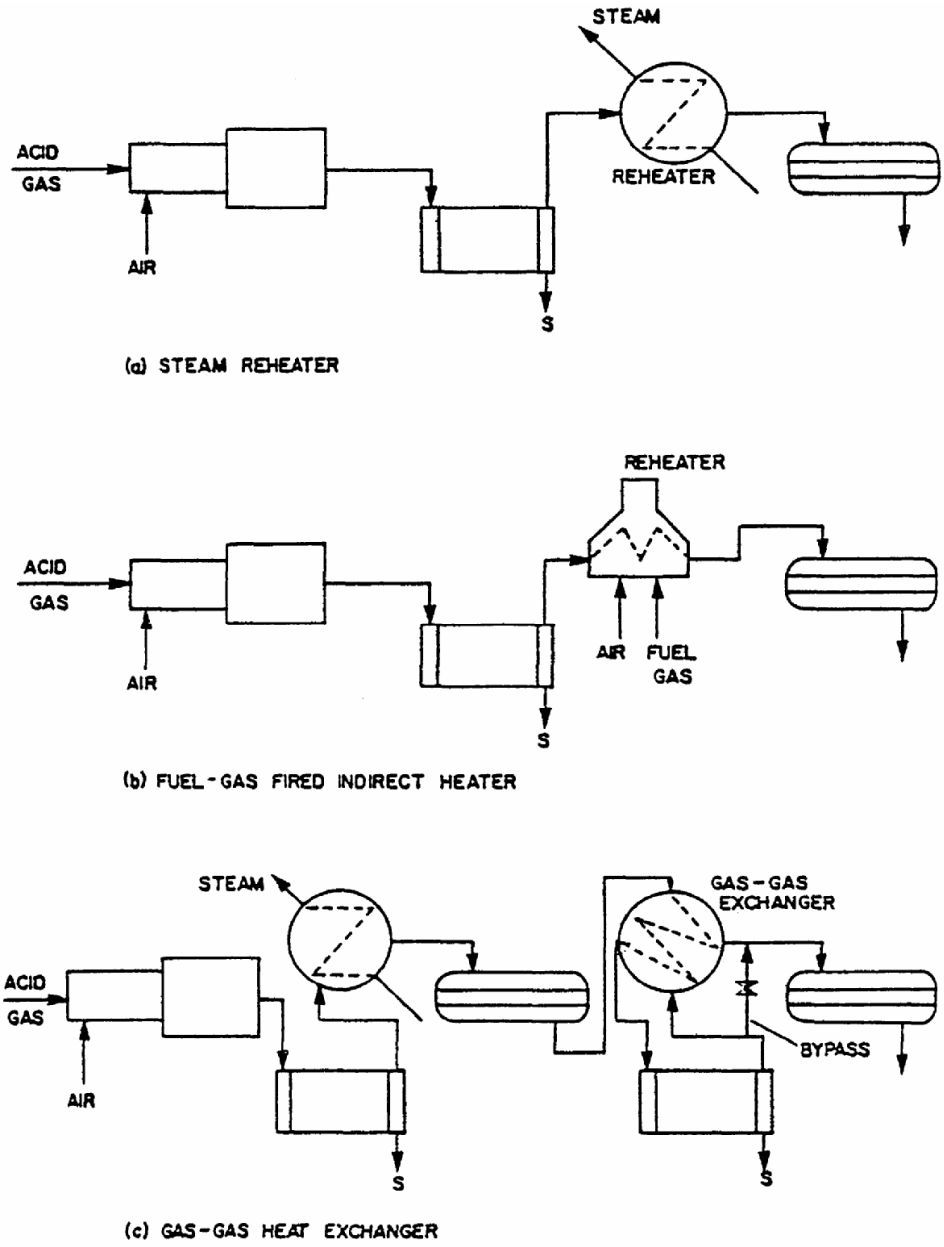


Figure 20.3: Indirect Heat Methods

To handle the various % of H_2S present various modification from the straight through process are used. These are:

H_2S in Acid Gas (Mole Percent)	Configuration
50 100 (Rich Feed)	Straight-through.
20 50 (Lean Feed)	Split-flow.
10 20 (Lean Feed)	Split-flow with preheating of feed streams.
< 10	Sulphur recycle or direct oxidation.

The ranges noted in the above table are not sharply defined - such factors as the presence of species other than H_2S and CO_2 in the acid gas, acid gas flow rate, stability of the acid gas composition and flow rate, past experiences of the designers, and economics have a large

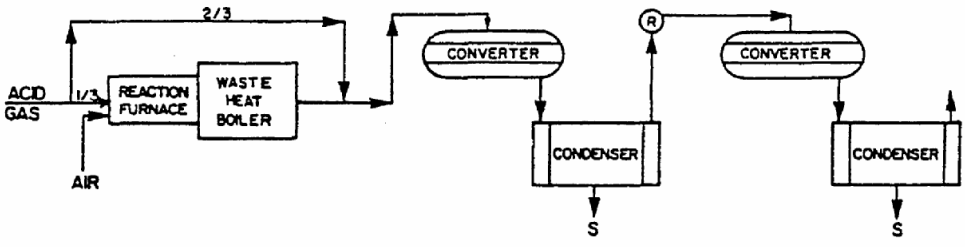
influence on the choice of process configuration. For example, the presence of large amounts of NH_3 in the acid gas may require a special burn strategy, as shown in Figure 20.5, or acid gases with very, very low H_2S contents might best be handled by a version of the original Claus process shown in the Table above.

Shown schematically in Figure 20.4 are three basic configurations that are in current usage:

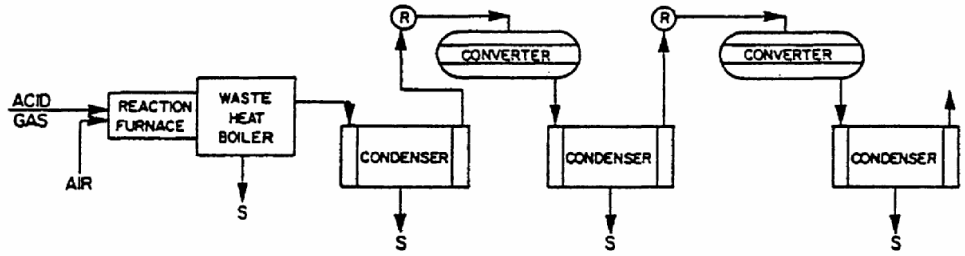
1. split-flow.
2. straight-through.
3. sulphur recycle.

Whatever the configuration, the string of vessels comprised of the reaction furnace, condensers, reheaters and converters is referred to as a sulphur recovery "train".

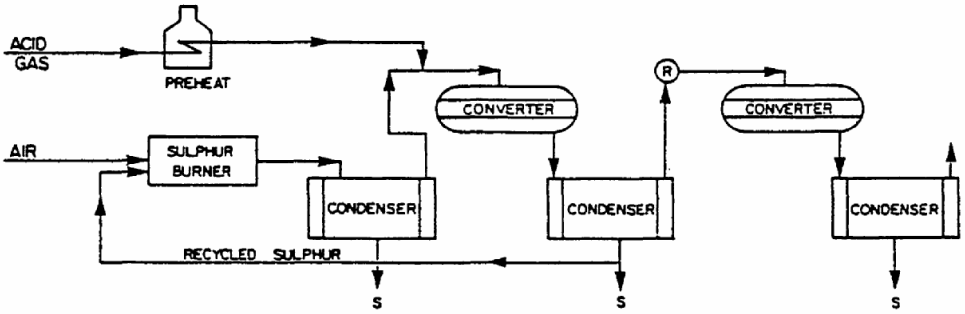
The differences among the configurations are in the methods used to produce the SO_2 prior to the first converter. These essential differences are noted below - the reaction furnace is discussed in detail later.



(a) Split - flow configuration



(b) Straight - through configuration



(c) Sulphur recycle configuration

Figure 20.4: Modified Claus Process Configurations

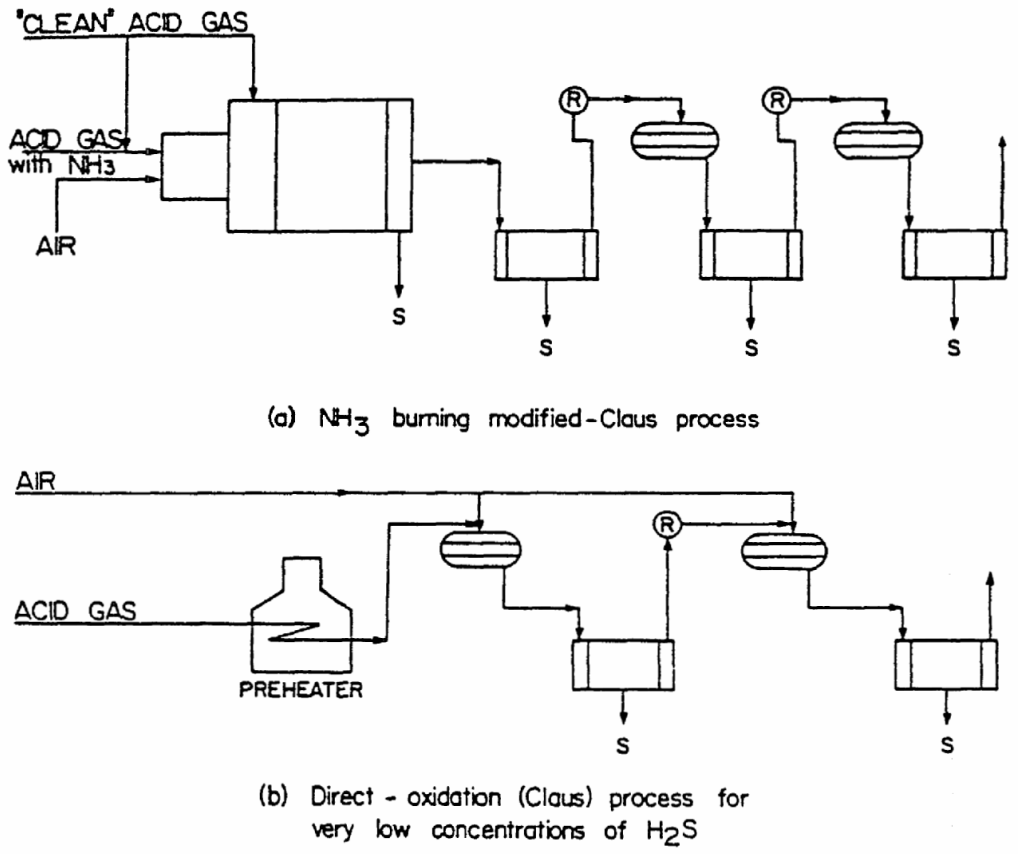


Figure 20.5: Special Versions of the Claus Process

In the first (split-flow), as little as 1/3 of the H_2S stream passes through the reaction furnace and boiler, the remaining fraction being mixed with the boiler exit gases just before the Claus oven (converter). In this configuration, significant elemental sulphur in the furnace is not produced in the furnace.

In the second (straight-through) configuration all of the H_2S goes through the reaction furnace where a significant part of the H_2S is formed into elemental sulphur leaving the rest of the H_2S and SO_2 to react in the converters.

In the reaction furnace, the primary requirement is a stable flame at a sufficiently high temperature to complete the reaction between O_2 and H_2S. Flame temperature is dependent on the H_2S content of the acid gas[†] and the fraction of acid gas bypassed around the flame. In those cases where the H_2S content of the acid gas is too low to achieve a sufficiently high temperature with a 2/3 bypass, acid gas preheating or the sulphur recycle is used or sometimes direct oxidation in which the acid gas is heated and fed directly along with the appropriate amount of air into the reactors. Please see further on in this chapter under "Furnace Conversion" for another discussion on this subject.

20.2.1 Equilibrium Conversion

[†] The "H_2S-containing stream" feeding the plant is referred to as the "acid gas" stream.

To understand what happens in a sulphur plant as it passes through the process, it might be well to consider the equilibrium conversion versus temperature. This data is given in 20.6. You will note the minimum conversion is near 1000º and that there are areas of catalytic conversion and of flame conversion. The flame conversion occurs in the reaction furnace and thus the warmer this can be run, the greater the direct conversion. The figure also shows the desirability in the catalytic step of keeping the temperature as low as possible.

20.2.2 Thermodynamic Data

The first intensive investigation of the thermodynamics of the Claus reaction was reported in 1953 by Gamson and Elkins. The results of their equilibrium calculations for pure H_2S are embodied in the lower solid curve in Figure 20.6. Their calculations were based upon the thermodynamic data available at the time, which for the sulphur vapour species were those derived in 1937 from 1909 measurements, and included only the species S_2, S_6 and S_8. Most predictions of modified-Claus process capability have been made from the same data set.

Since 1953, the existence of the additional sulphur vapour species S_1, S_3, S_4, S_5 and S_7 has been confirmed by a number of workers and the composition of saturated sulphur vapour measured. The vapour pressure measurements of West and Menzies have been extended down to ambient temperature (77ºF) and up to the critical point (190ºF). However, the thermodynamic data sets derived by various workers from these, and other, data were not in agreement. This led to scepticism of the data and they were therefore usually ignored for Claus process calculations.

While recognizing the presence of species other than S_2, S_6 and many workers assumed that the effect on the modified-Claus equilibrium calculations was insignificant and continued to include only S_2, S_6 and S_8 with old or new data. In some cases, all species were considered for sulphur dewpoint calculations, but not for reaction equilibrium.

After carrying out calculations for both cases Maadah and Maddox concluded that there was little difference in the calculated recovery efficiencies whether all sulphur species were considered or only the species S_2, S_6 and S_8 were considered. For the pure H_2S case, their cumulative efficiencies calculated with all sulphur species after the first, second and third converters were higher by 0.53, 0.28 and 0.13 percent, respectively. The converter and condenser temperatures were not stated.

The current trend is to consider the presence of all sulphur vapour species for modified-Claus equilibrium calculations. This work continues this trend, but with an approach that is different than most others.

Thermodynamic data for the sulphur vapour species were developed from the combination of information obtained from the literature and measurements made on the modified-Claus reaction, both in actual sulphur recovery plants and in the laboratory. The resulting sulphur vapour species thermodynamic data set predicts equilibrium conversion efficiencies that are consistent with measured values. Thermodynamic data for all other species were obtained from the literature.

All three curves in Figure 20.6 were calculated without sulphur removal from the system, and both data sets indicate that below 540ºF, liquid sulphur should be formed. If the calculation procedure does not allow liquid sulphur formation, a non-physical condition results and the calculated equilibrium conversions are too low.

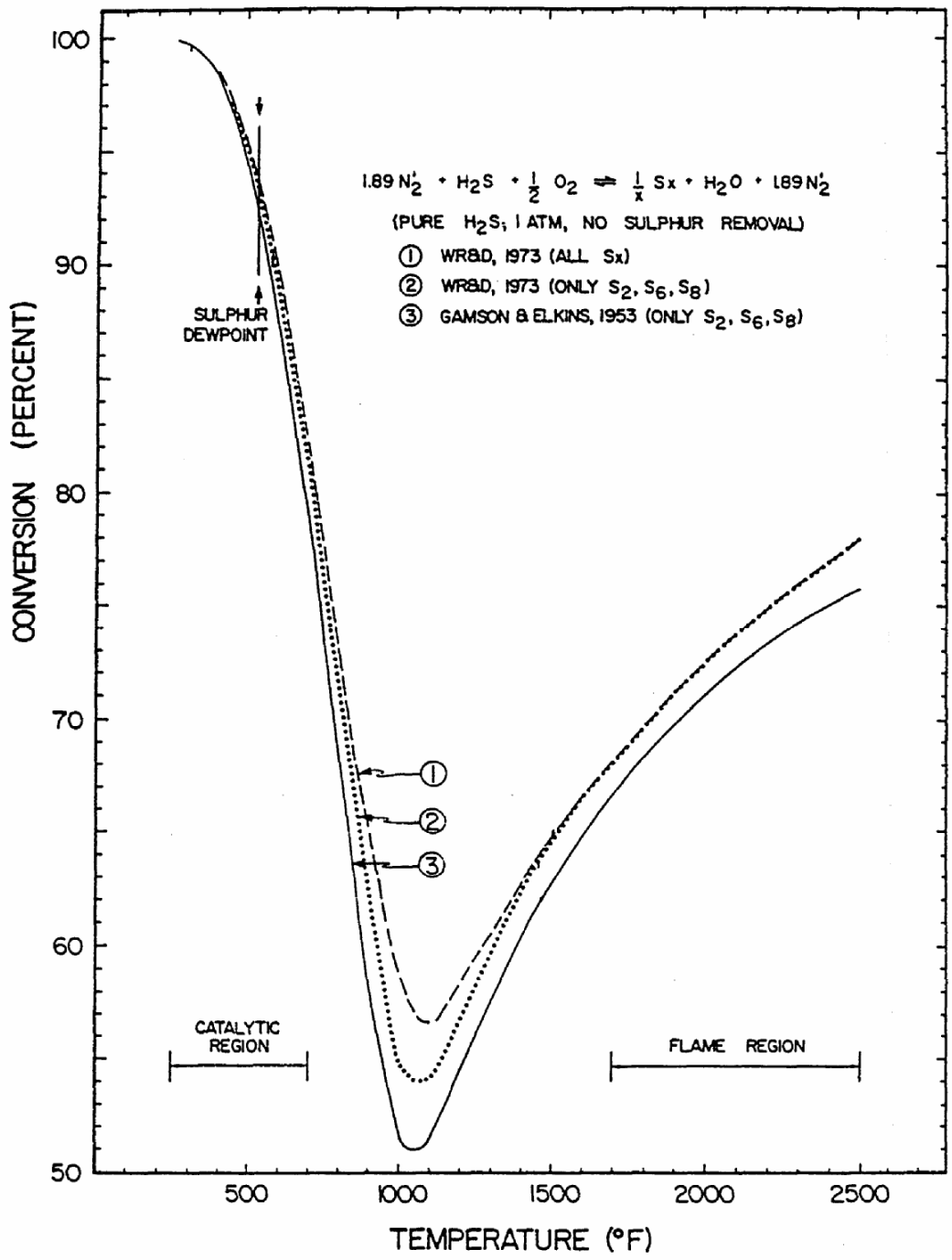


Figure 20.6: Equilibrium Conversion of H_2S to Sulphur

20.2.3 Reaction Furnace

The usual first and second process units of a modern sulphur train are a free flame reaction chamber and a heat exchanger. They are very often combined into a single vessel and are commonly referred to as the reaction furnace and waste heat boiler. It was the addition of these units to the original Claus process during the 1930's that resulted in the present day modified-Claus process.

The functions of the reaction furnace and waste heat boiler are to:

- burn 1/3 of the H_2S (with air) to SO_2 for subsequent reaction with the remaining 2/3 of the H_2S in the converters.
- produce elemental sulphur directly by the partial oxidation of H_2S.
- destroy any contaminants in the H_2S-feed stream.
- cool the exit gas stream to the lower temperatures that favour the H_2S-SO_2 reaction.

In the first versions the 1/3-2/3 split was obtained by physically dividing the acid gas stream and bypassing the 2/3 portion around the flame and boiler. It was mixed with the boiler outlet stream just ahead of a catalytic converter. This configuration fulfilled the first and last functions, and part of the third.

It was recognized that a significant quantity of sulphur would be produced if all the acid gas passed through the flame with the reaction being controlled by the amount of air admitted. Sulphur removal in a condenser just before the first converter would enhance sulphur formation within the converter. Moreover, the temperature rise across the first converter is substantially reduced. Further, the acid-gas contaminants are destroyed and heat recovery in the boiler is maximized by this straight-through configuration.

20.2.3.1 Furnace Chemistry

Complete combustion of pure H_2S with pure O_2 will yield an adiabatic temperature of about 8600ºF. However, if only 1/3 of the H_2S, SO_2 and H_2S, the temperature will be about 4,700ºF; further reaction to completion with only S_2 and H_2O as products will yield a final temperature of 4,480ºF. If air is used instead of pure O_2 the energy consumed in heating the N_2 reduces the temperatures to 3,380ºF, 2,630ºF and 2,460ºF, respectively, for these conditions. All the above temperatures were calculated with an assumed initial temperature of 100ºF.

The actual temperature reached depends upon the types and quantities of other acid gas constituents such as CO_2, H_2O, hydrocarbons and residual sweetening agents. Combustibles tend to increase the temperature, for obvious reasons, while the so-called "inerts" tend to decrease the temperature by the simple dilution effect and as reactants in endothermic reactions. The reported range of temperature is 1,400ºF to 2,300ºF, the lower temperatures being obtained in lean feed plants. The majority of plants operate in the 1,700ºF to 2,200ºF range.

Basic chemical reactions occurring in the reaction furnace are summarized in Table 20.2. The first three are the usual modified-Claus and Claus reactions, and the next four show some of the additional reactions that very likely occur. The last four illustrate the combustion of some hydrocarbon contaminants that are frequently found in the acid gas streams. The reaction between O_2 and H_2S is very fast, and usually goes to completion in the furnace.

Table 20.2: Modified-Claus Process Reaction Furnace

	Reaction (gram-moles)	F^* (kcal)		H^{**} (kcal)	
		1700°F	2200°F	1700°F	2200°F
1.	$3H_2S + \frac{3}{2}O_2 \rightarrow H_2O + 2H_2S$	-101.5	-96.2	-124.6	-124.5
2.	$2H_2S + SO_2 \rightarrow \frac{3}{2}S_2 + 2H_2O$	-6.3	-10.1	10.1	9.9

3.	$3H_2S + \frac{3}{2}O_2 \rightarrow \frac{3}{2}S_2 + 3H_2O$	-107.8	-106.3	-114.0	-114.6
4.	$H_2S + \frac{1}{2}O_2 \rightarrow H_2O + S_1$	-1.3	-4.8	13.9	13.8
5.	$S_1 + O_2 \rightarrow SO_2$	-100.2	-91.4	-138.4	-138.5
6.	$2S_1 \rightarrow S_2$	-69.3	-61.3	-103.3	-104.2
7.	$S_2 + 2O_2 \rightarrow 2SO_2$	-131.2	-121.5	-173.0	-172.8
8.	$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	-191.1	-191.0	-191.2	-192.2
9.	$C_2H_6 + \frac{7}{2}O_2 \rightarrow 2CO_2 + 3H_3O$	-355.9	-359.0	-341.7	-342.8
10.	$C_6H_6{}^{a} + \frac{7}{2}O_2 \rightarrow 6C + 3H_2$	-71.8	-85.1	-144.6	-142.4
11.	$C_7H_8{}^{b} \rightarrow 7C + 4H_2$	-219.4	-245.1	-80.1	-145.6

* Change in Gibbs free energy for reaction as written. Negative change indicates spontaneous reaction possible.

** Heat of reaction as written. Negative ΔH indicates exothermic.

[a] benzene, gaseous

[b] toluene, gaseous

It has been pointed out that the complexity of the furnace reactions is much greater than these equations imply. Indeed, there is very likely a myriad of free-radical intermediates involved. Nonetheless, the net reactions like those in Table 20.2 are sufficient for most practical applications.

A major failing of the listed equations is that they do not account for the H_2, CO, COS or CS_2 observed in operating sulphur plants. There are many side reactions which can occur. Please see Paskell for further details.

20.2.3.2 Furnace Conversion

The primary functions of the reaction furnace and waste heat boiler are to generate SO_2 for later reaction with H_2S in the downstream converters, and to recover what would otherwise be lost energy. Therefore, any conversion to elemental sulphur achieved in the furnace is a fringe benefit. If the sulphur is removed, the load on the converters is reduced in terms of throughput, heat dissipation and sulphur dewpoint. All these effects lead to enhanced conversion.

For the straight-through mode, the furnace conversion is given by the high temperature portions of the equilibrium curves in Figure 20.6. It is primarily a function of temperature although it is also affected by total system pressure. Above 1,700ºF, there is negligible difference between the conversions achievable by rich and lean feed plants. For one atmosphere pressure, the equilibrium conversion increases from about 68 percent at 1,700ºF to about 75 percent near 2,300ºF. In this range, a change in total pressure from 1/2 to atmospheres will result in about a 5 percent decrease in the conversion.

The furnace conversion of a split-flow plant is dependent upon the quantity of acid gas bypassed around the flame for two reasons. Firstly, as soon as any acid gas is bypassed without also

bypassing a like fraction of process air, the furnace is forced into an excess air condition which results in lower equilibrium conversion. Secondly, any H_2S that bypasses a unit is not available for conversion in that unit. Furnace conversion as a function of bypass fraction is shown in Figure 20.7.

At 2/3 bypass, the conversion is necessarily zero, since all the H_2S passing through the flame is oxidized to $S0_2$. At zero bypass, the conversion is limited by the lower flame temperature. At intermediate points, the conversion is the net result of the interplay among the excess air condition, increased flame temperature and fraction of the acid gas bypassed.

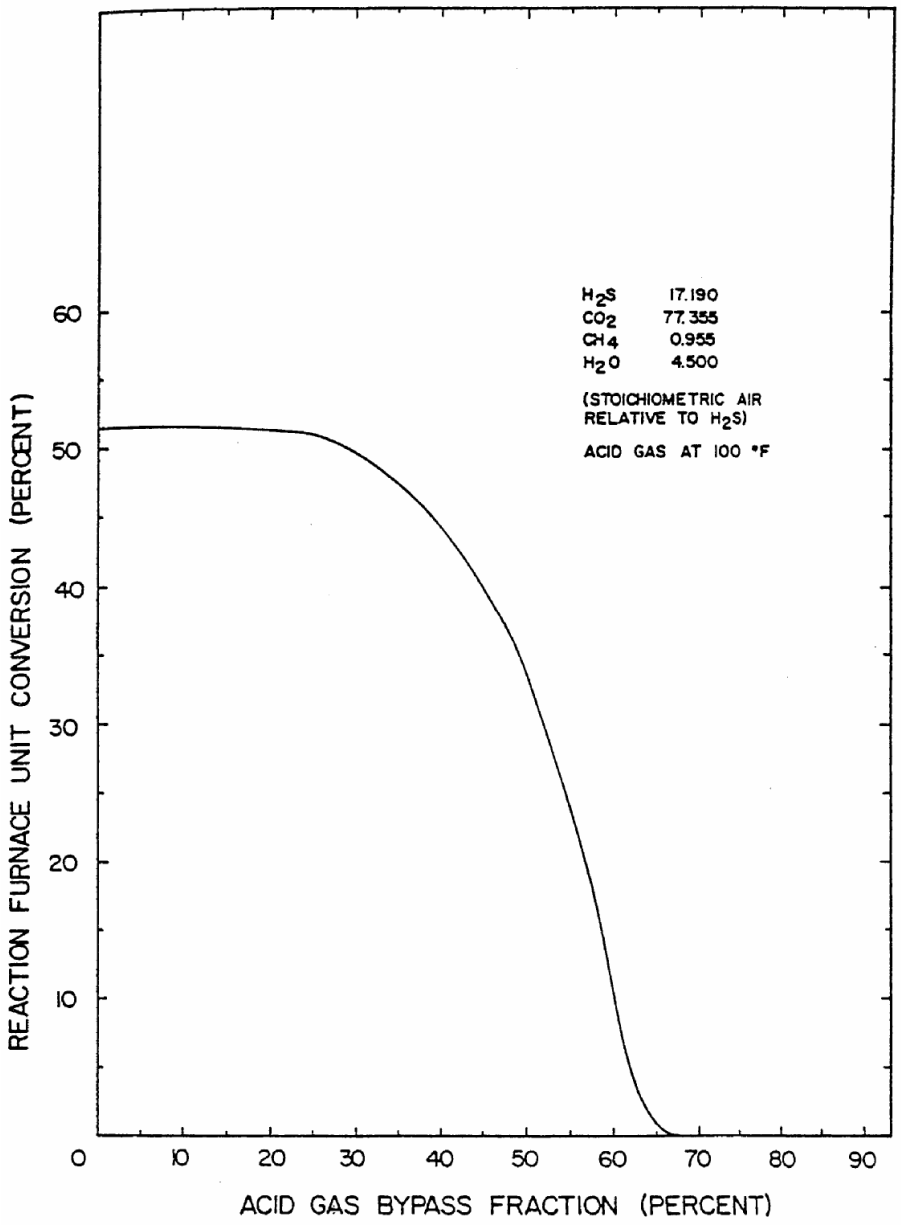


Figure 20.7: Reaction Furnace Conversion Versus Acid Gas Bypass Fraction

The question of how much acid gas to bypass is not always easily answered. If too much is bypassed, the flame temperature will be too low. For very lean feeds, the minimum and maximum

bypass fractions are often the same, which can result in operational problems of striking a balance between a reasonably stable flame and preventing O_2 from reaching the first converter.

From Figure 20.7, it would seem that conversion efficiency is maximized with minimum bypass fraction. If equilibrium is always achieved, this would be true. Since reactions are kinetically limited in most lean feed sulphur plant furnaces, the ideal bypass fraction is best determined empirically after the plant is in operation.

Field conversion measurements at a plant with a 33 percent H_2S feed at various acid gas bypass fractions are shown in Figure 20.8. Since the relative flows of acid gas to the flame and bypass were not metered, the split was determined by material balance calculations at the highest bypass fraction. At intermediate points the furnace temperature was used as a guide, and for the equilibrium calculations the fractions were determined by linear interpolation.

As the bypass fraction increased, conversion improved as a sequence of the higher temperature and longer residence time. The flame quality and stability also improved. These data clearly show that at the lower temperatures and short residence times, furnace reactions were kinetically limited. For this plant, the best bypass fraction is very near the maximum practical fraction. Bypassing a larger fraction increases the danger of breakthrough to the first converter.

Earlier, it was shown that even with zero furnace conversion in the reaction furnace the thermodynamic capability of the lean feed plant was only slightly less than for a rich feed plant having 45 percent recovery before the first converter. For this reason, maximization of furnace conversion should not be the criterion for selecting the acid gas bypass fraction. The criteria should be stable operations and a high temperature with a long residence time.

Kinetic limitation of reaction furnace conversion is a general observation for lean feed plants. This is shown in Figure 20.9. While many of the rich feed plants achieve equilibrium conversions in the furnace, the lean feed plants in general do not.

Reaction 20-4

$$H_2S + \frac{3}{2}O_2 \rightarrow H_2O + SO_2$$

$2H_2S + SO_2 \rightarrow 2H_2O + 3S$, or summing

$$3H_2 + \frac{3}{2}O_2 \rightarrow 3H_2 + 3S$$

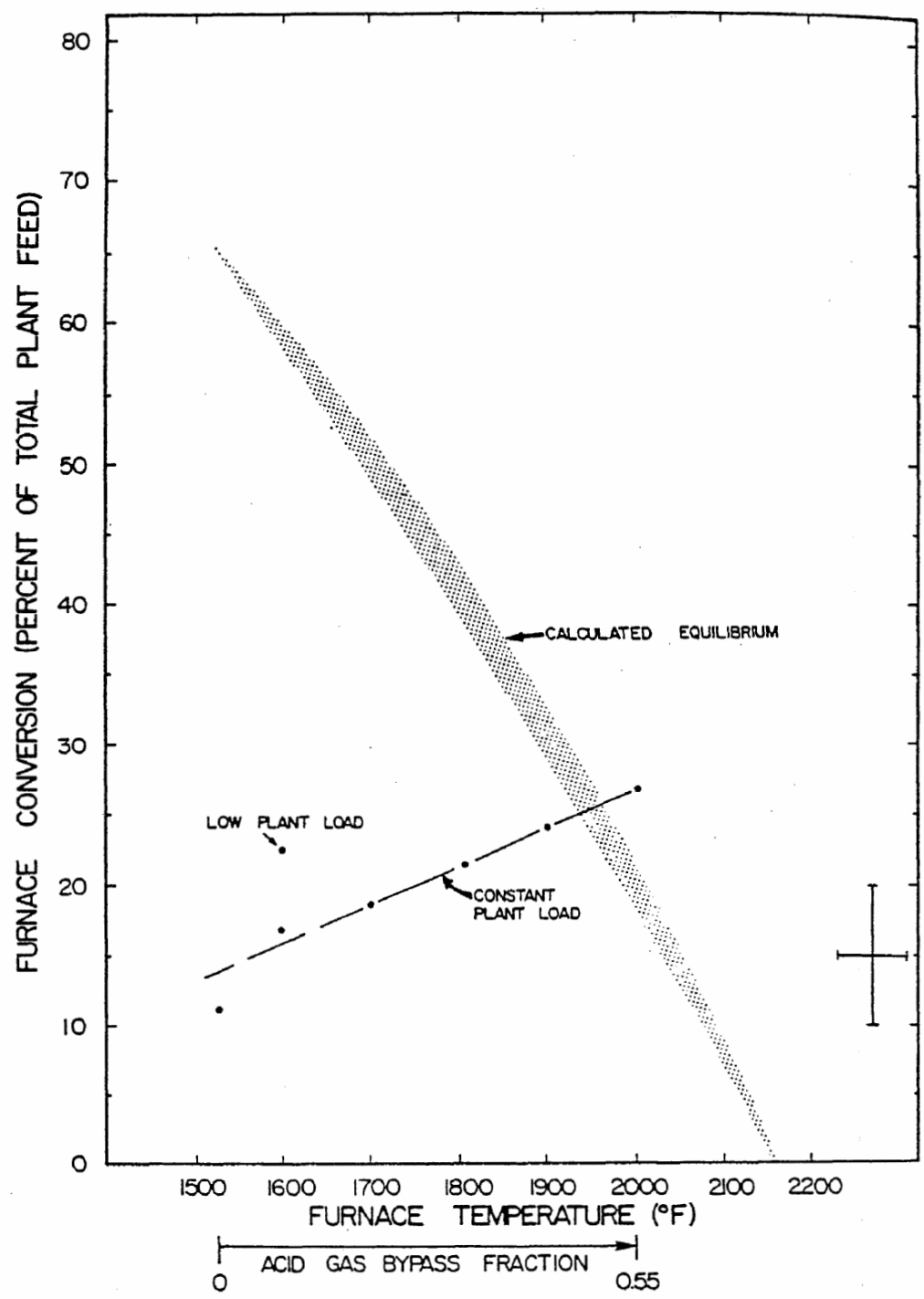


Figure 20.8: Furnace Conversion Versus Acid Gas Bypass Fraction (Measured)

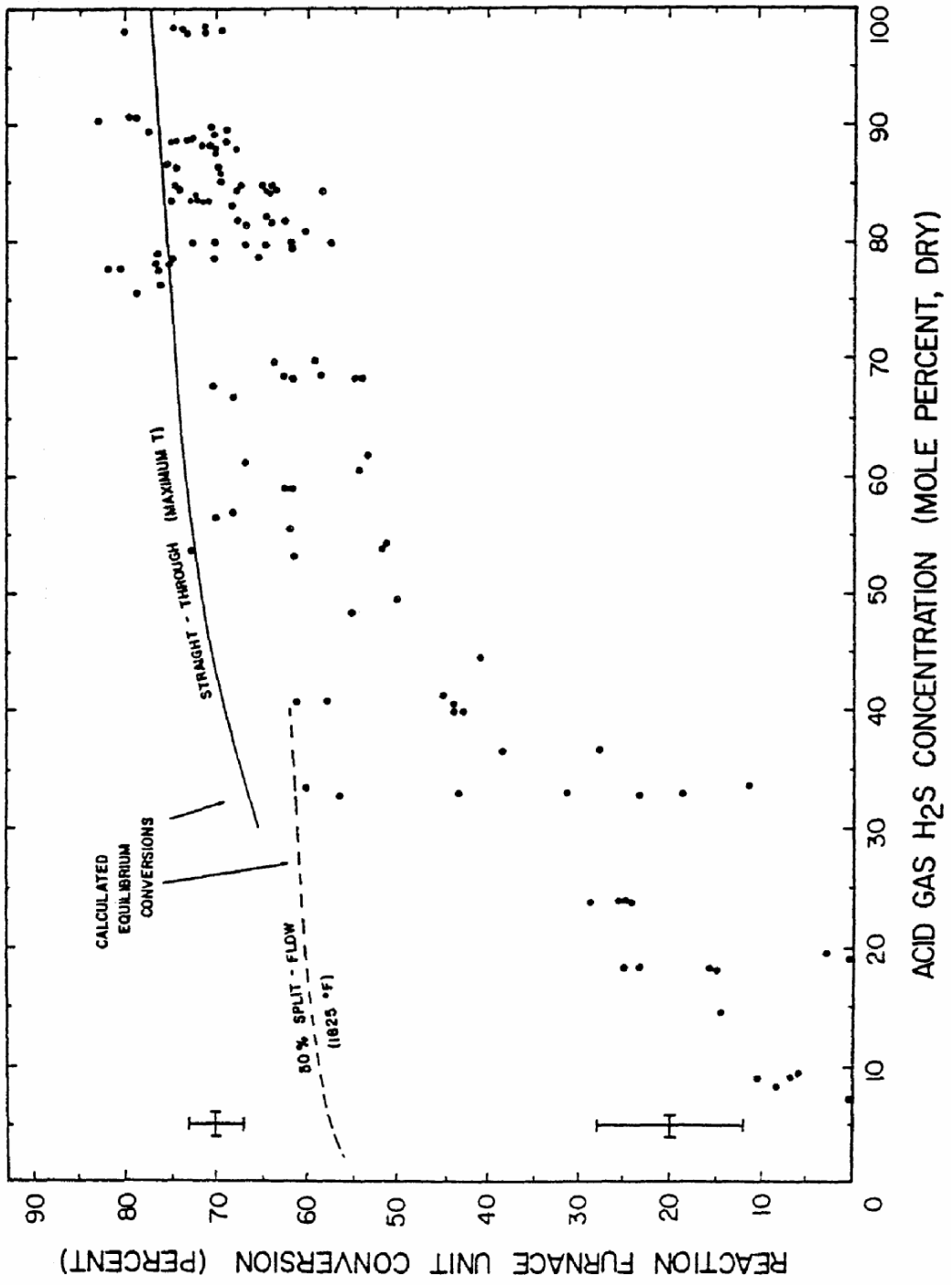


Figure 20.9: Measured Reaction Furnace Unit Conversions

20.2.3.3 Process and Mechanical Considerations

The normal design residence time in the furnace is 0.5 seconds.

The refractory design is very important as it must be such that it can stand high temperatures (2,000ºF) and have good insulating properties so that the steel shell is below 650ºF so that softening of same does not take place. Thus both high alumina and insulation bricks are used. Also internal checker walls are often provided for the combustion reaction to take place.

20.2.3.4 Gas Cooling & Sulphur Condensation

Gases leaving the combustion furnace at 2,000ºF to 2,500ºF are most often cooled by generating saturated steam in a waste heat boiler of the fire-tube type. Hot gases pass through the tubes with boiling water on the shell side where steam pressures between 20 and 600 psig can be successfully maintained. Sulphur is not normally condensed in the waste heat boiler because the outlet temperature (500ºF to 650ºF) is above the sulphur dewpoint temperature for gases leaving the furnace.

For straight-through type units, furnace-produced sulphur is roughly 50%-65% of the total production. The first sulphur condenser downstream of the waste heat boiler recovers sulphur by cooling gases to 300ºF-400ºF. Cooling is usually accomplished by steam generation at steam pressures between 20 and 80 psig. Gas outlet temperature from the sulphur condensers is limited to about 20ºF minimum above temperature.

If low-pressure steam (20-80 psig) is not needed in the plant, another procedure occasionally used is a "closed" water boiling system where steam is generated in sulphur condensation at 20 psig, condensed at the condensate returned to the "boiler" (sulphur condenser) as feed water, thereby "closing the loop". This cooling method minimizes the amount of make-up water required for the system.

20.2.4 Waste Heat Boiler

Boilers other than large fire-tube use smaller tubes which are exposed to hot combustion gases. Consequently, these tubes must be protected to ensure satisfactory life. Additionally, the inlet tubesheet is protected from direct contact with hot gases by use of about 3 inches of refractory.

The inlet end of each tube is shielded by inserting into the tube a ceramic ferrule that extends about 6 inches in front of the inlet tubesheet to prevent hot gas from contacting the tube directly at the critical junction of tube and tubesheet. Use of ceramic ferrules is imperative if reasonable tube life is to be expected. For best results, these ceramic ferrules are wrapped with a thin layer of insulation (1/8-3/8 inch thick).

Tube size ranges from 2 to 6 inches for most boilers. Generally, welded (ERW) tubes are installed, but occasionally seamless tubes are used. Tube spacing is based on a minimum ligament dimension of 0.75-1.0 inch. Design mass velocity ranges between 1.0 and 8.0 lb/sec/ft^2 based on allowable tube-side pressure drop.

Tubesheet design (ASME code, Section I) for these vessels is greatly influenced by thermal/mechanical stress at the tube attachments. Tubesheets are usually thin (3/4 to 1 1/2 inch), resulting in acceptable temperature profiles at tube welds, tubesheet ligaments between tubes, and for portions not backed by water. Deflection that occurs during operation must be considered in order to reduce additional stress transmitted to the tube attachment. Boilers that are fully flooded have tube patterns arranged to account for bending stress at the outer boundaries of the tubesheets, where as kettle-type boilers have stayrods to stiffen portions of tubesheets not supplied by tubes. Where possible, boilers shall be provided with bolted inspection ports at the end and/or beginning of each pass.

Proper process design of the boiler requires that heat exchange calculations include effects of radiation transfer as well as normal convection transfer. If radiation is ignored in calculating surface area requirements, the Claus boiler is usually appreciably over-sized. This can lead to operating problems because a waste heat boiler that is too large can cause premature sulphur condensation, improper reheat, viscous sulphur Formation, etc.

20.2.5 Converters

20.2.5.1 Design

These are generally horizontal vessels with relatively thin long catalyst beds. Figure 20.10 shows a typical vessel in this service.

These are sized for a space velocity between 700 to 1,100 reciprocal hours or:

Equation 20-1

$$Flow/catalyst\ volume = \frac{CF/hr}{CF} = 700 - 1{,}100$$

The flow is the total process gas and is measured at standard conditions. This can also be expressed as 2-3 moles of process gas per cu. ft. of catalyst.

The lower the entrance temperature to the converter the better conversion as indicated by the equilibrium graph (Figure 20.6). The normal entrance temperature is 450ºF or higher in the first converter (so as to help COS and CS_2 hydrolize) and lower to the converters as the sulphur dewpoint temperature is just below 400ºF. The diameter of the vessel is usually set by pressure drop calculations. The pressure drop should be minimized.

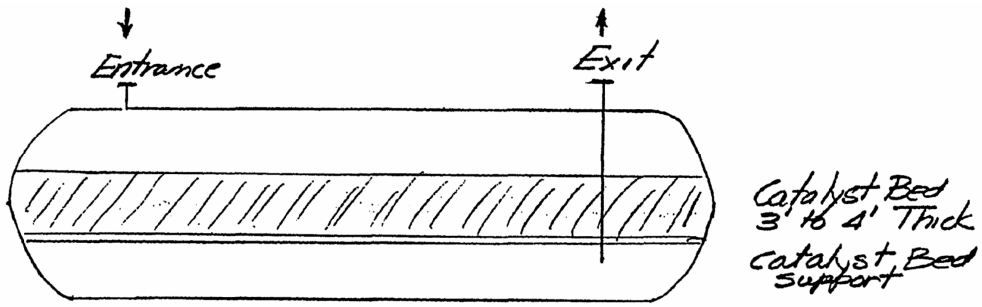


Figure 20.10: Typical Converter

20.2.5.2 Modified-Claus Catalysts

A complete discussion of modified-Claus reaction catalysts is beyond the scope of the present chapter. Research on this topic has been carried out by many workers and further work is continuing. The primary purposes of the research are to define the desirable catalyst properties and to determine reaction rates so that catalysts may e selected objectively and converters sized properly. Deactivation mechanisms have also been studied with the view toward developing methods of prolonging catalyst life. Although not complete, the research has been fruitful in both areas. A significant finding from both research and plant experiences is that catalyst life is highly dependent upon operating practices. Useful life spans of the catalysts vary widely from plant to plant, from one to thirteen years.

Desirable catalyst properties include: high catalytic activity, ready availability, low cost, high resistance to deactivation and aging mechanisms, highly mechanical strength and easy handling,

low resistance to gas flow, and high resistance to attrition. In general, activated bauxites and aluminas satisfy these requirements and are the catalysts generally used in the modified-Claus process. Primarily alumina is now used. Sufficient operating experience has been gained with these catalysts to show that, when new, they are entirely satisfactory.

Activity of a catalyst in general is dependent upon both physical and chemical properties: composition, stereochemical structure, particle size, porosity and surface area. The blocking or destruction of any one or all of these properties will reduce the catalytic activity.

In addition to the activity, the factors that affect overall catalyst bed performance include: temperature, pressure, concentrations of reactants and products, and contact time.

The primary concern is that the activity of the catalyst is sufficient to promote the reaction to equilibrium within the converter. Fresh bauxite catalysts generally satisfy this criterion as illustrated by the upper reaction profile in Figure 20.11. Equilibrium conversion is reached within a foot of bed depth. For given conditions, the activity of alumina is usually much greater than that of bauxite.

In laboratory studies of modified-Claus reaction kinetics such as the above, the measured conversion efficiencies are often much higher than predicted by equilibrium calculations. Since a catalyst can affect only the reaction rate, and the equilibrium extent of reaction, this implies an error in the equilibrium calculations, the measurements or in both. As pointed out previously, the comparison of measured quantities with equilibrium predictions requires that the equilibrium calculations be carried out for the exact experimental conditions that existed for the measurements. The prime experimental condition is that equilibrium calculations be carried out for the exact experimental conditions that existed for the measurements. The prime experimental condition is that equilibrium be established in the system before the measurements are made.

For modified-Claus plants, the activity of fresh catalyst is not a problem bauxites and aluminas are readily available. Deterioration of catalytic activity is the major problem, and this appears to be related more to feed contaminants and operational problems than to intrinsic properties of the catalyst material.

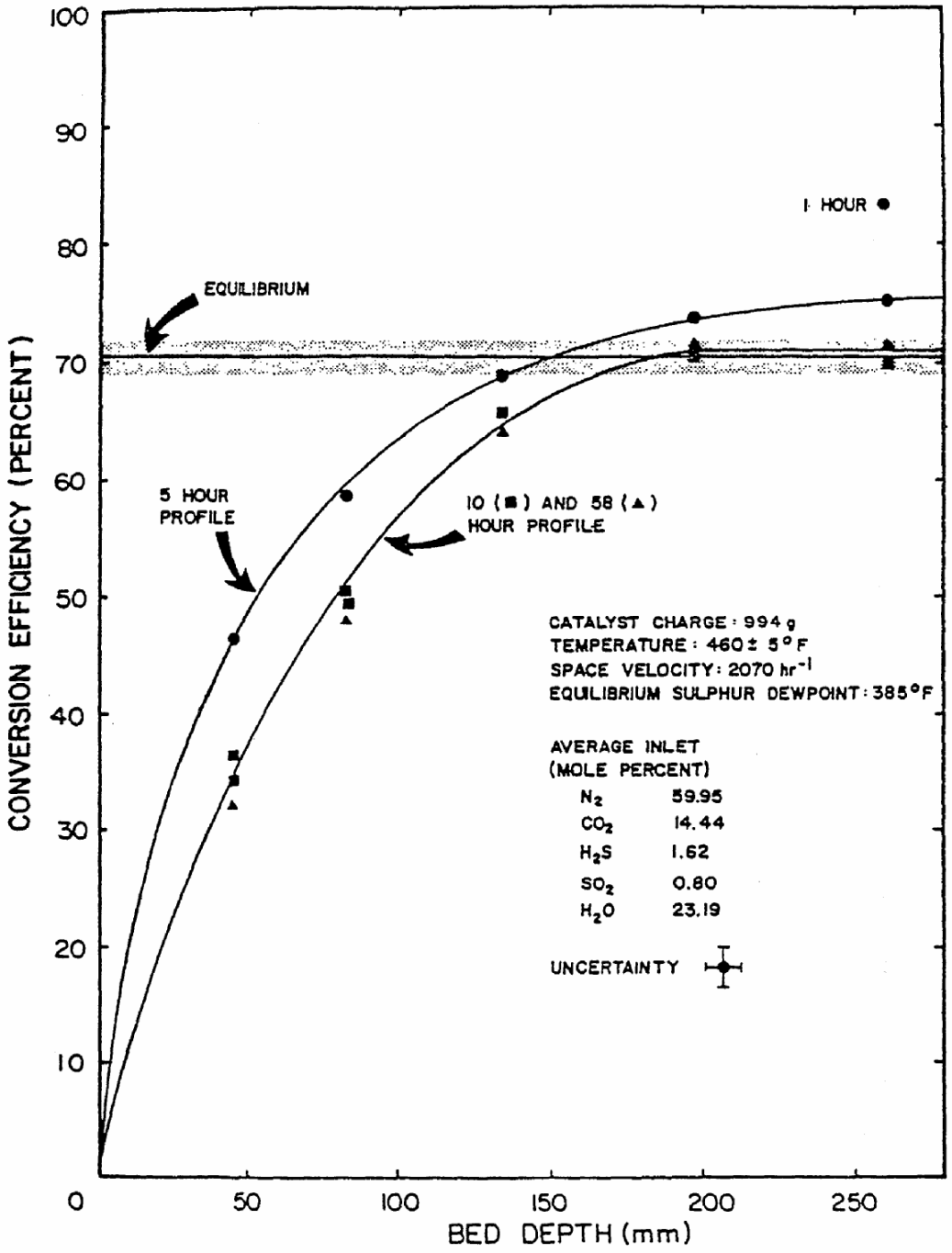


Figure 20.11: Modified-Claus Reaction Over Bauxite After Specified Hours of Continuous Operation

20.2.5.3 Deactivation and Catalyst Management

In most sulphur plants, the catalyst bed depth is 3 to 4 feet, and it is generally observed with fresh catalyst that the reactions reach equilibrium within the first 6 to 12 inches. As deactivation proceeds, the overall conversion by the converter is not affected until the full bed depth is insufficient for the reactions to reach equilibrium. From this point on, the conversion by the bed is kinetically limited. Its occurrence is signaled by a decrease in the temperature rise in the following

converter. If the last converter becomes kinetically limited an immediate decrease in overall recovery efficiency will be observed.

The catalyst deactivation mechanisms can be grouped into two general categories. The first category contains those mechanisms that involve changes in the intrinsic structural properties, for example, surface area loss due to thermal aging, phase changes, sintering, and attrition. These mechanisms are non-reversible and cannot be entirely eliminated. For normal modified-Claus operating conditions, however, they are very slow. More damage is done in this area by excessive temperatures during plant startups and shutdowns and the so-called regenerative burnoffs of catalyst deposits than by normal plant operations.

The second category involves external agents which by chemical reaction with or simple deposition on the catalyst block access of the reactants to the active sites. Major mechanisms in this category are: sulfation, sulphur, sooty and glassy carbons, salts and pipe corrosion products.

In general, deactivation by these mechanisms is reversible, although for carbons and corrosion products the regeneration procedure may severely shorten the catalyst life by the previously discussed mechanism. If chronic problems are encountered with these types of deactivation, the only effective solution is to eliminate the cause. This may simply require a change in the way certain streams, such as the lean feed acid gas bypass are handled or a change of the inline burner fuel. Or, it may require the removal of contaminants from the feed streams, for example, removing NH_3, reducing the frequency of foaming in the sweetening process, or simply reducing sweetening agent and hydrocarbon carryover.

Deactivation by carbon is highly dependent upon the form of the carbon deposit. On the one hand, sooty carbon, such as that formed by CS_2 cracking or substoichiometric combustion of light hydrocarbons, does not significantly deactivate the catalyst. Indeed, carbon in the form of activated charcoal is used as the catalyst in the original Sulfreen process. Sooty carbon may become a problem if sufficient quantities are deposited to cause excessive pressured or contamination for the sulphur. On the other hand, if the carbon deposit is of the glassy variety, or tars, deactivation can be rapid and severe. This type of material encases the catalyst pellets in a "protective" coating that completely blocks access of the reactants to the active sites.

Adsorption of sulphur vapour by the catalyst does reduce catalyst activity, but obviously it is unavoidable. The quantity of sulphur taken up by the catalyst is dependent upon temperature, being highest at low temperatures. The adsorption is reversible and partially accounts for initial high recovery efficiencies following plant startup and the higher emissions at shutdown.

The deposition of liquid sulphur, however, will completely deactivate the catalyst. Fortunately, this deactivation is completely reversible by simply raising the converter temperature.

For converters operated at low temperature, but above the dewpoint (eg. third and fourth converters), the build-up of sulphur within the catalyst pores by adsorption and capillary condensation may block access to the active sites, and kinetically limit the conversion. This type of deactivation can be remedied online by temporarily raising the converter inlet temperature by 25ºF to 50ºF to drive off the sulphur. After 24 to 36 hours, the inlet temperature is slowly reduced to the normal operating temperature. This method, the "heat soak", is being widely practiced in the industry. It is far superior to the old-style burnout procedure.

Sulfation at the active sites is considered the major deactivator of modified-Claus catalysts. Unfortunately, a certain amount of sulfation is inevitable because of SO_2 chemisorption. More seriously, the sulfate levels can be rapidly increased by SO_3 formation in, or O_2 breakthrough from, the reaction furnace or inline burners, or through combustion of adsorbed sulphur during startup or shutdown.

The breakthrough of O_2 from inline burners is harmful from other aspects as well. It increases the temperature rise across the converter which misleads the operator into believing that the converter is performing well.

It reduces conversion because of the higher temperature. If inline burners must be used, they should be the most efficient available, be well maintained, the air to fuel ratio doubly checked and their operation carefully monitored, particularly during startup and shutdown.

Sulfation is reversible, and a number of reducing agents besides simple thermal decomposition can be used. The presence of some of these agents (H_2S, H_20) in the process stream makes possible online rejuvenation of the catalyst. Before reducing the temperature after the previously described heat soak used to remove adsorbed sulphur, excess H_2S is forced in the process streams by deficient air operation of the plant ($[H_2S]/[S0_2]$ = 2.5 to 10)[¥]. After 24 to 36 hours, the plant is returned to normal operating ratios and temperatures.

It is recommended that the heat soak and rejuvenation procedures be carried out on a regular basis, perhaps as often as once a month for the heat soak and 3 to 4 times per year for the rejuvenation.

The results of field measurements, confirmed by thermodynamic equilibria calculations, show that very high sulphur recovery efficiencies can be obtained with the modified-Claus process using fresh catalyst. However, in order to sustain such high efficiencies, these persistent catalyst deactivation problems must be solved. It is essential that research on these problems be continued.

20.2.5.4 Temperature Profiles

In a sulphur plant it is generally not possible to measure reaction profiles in the converters like those presented from laboratory reactors. Process stream sampling points are usually present only on the converter inlet and outlet process piping. Even if sampling probes were to be installed within the catalyst bed, the number required to adequately characterize the reaction distribution in such large vessels would be prohibitive in terms of installation costs, hindrance in catalyst change out, plugging problems and time and costs for sampling, analyzing and reducing data to obtain the profiles. Therefore, for operating sulphur plants, measurement of conversion efficiencies are usually limited to converter inlets and outlets only. Unfortunately, kinetics of reactions within the converter cannot be deduced from these data alone.

During construction it is relatively inexpensive to install thermocouples within the catalyst bed - the more, the better. The monitoring of the thermocouple readings to obtain temperature profiles within the catalyst beds is simple and fast - often the temperatures are recorded continuously on a strip-chart.

While there is no fixed relationship between conversion efficiency and converter inlet/outlet temperature differential, the shape of the temperature profile within the bed can provide valuable information about the condition of the catalyst. This is particularly true if the change in shape of the profile with time is monitored. Because O_2 breakthrough to the catalyst bed drastically changes the temperature profile in the bed, temperature profile analysis must be tempered with caution for first converters in lean feed plants and any other converters preceded by an inline burner.

Previously, it is shown that the temperature rise in the catalyst bed is very nearly a linear function of the extent of reaction. Therefore, the temperature profile is the same shape as the reaction profile.

Four temperature profile shapes are presented in Figure 20.12. Curve ‘A’ represented the shape observed for fresh catalyst - the reactions are usually near completion within the first 6 to 12 inches

[¥] High emissions during this phase may require a reduction of the plant throughput.

of bed depth. Curve ‘B’ type occurs when the catalyst deactivation is uniform within the bed such as with thermal aging. Deactivation of the catalyst from the top down, for example, by deposition of liquid sulphur or solid particulate, or crusting will result in a temperature profile like Curve ‘C’. Curve ‘D’ represents kinetic limitation from and advanced case of uniform deactivation.

The monitoring of converter temperature profile shape coupled with periodic measurement of conversion across the converter and sulphur production from the next condenser can be an effective tool for projecting catalyst life expectancy. Further, careful analysis of the profile shape changes can provide clues to the nature of the deactivation mechanism.

Temperature profile analysis for following catalyst condition is usually limited to first and second converters in which there are substantial temperature rises. However, irregularities in the profiles of any converter can also be used to judge other aspects of converter operation. These include the detection of poor gas flow patterns resulting from poor mixing of reheat streams with the main process stream or from the formation of a carbon/sulphur crust on top of parts of the bed, detection of liquid sulphur deposition, and, of course, the detection of sulphur fires in the bed during startup or shutdown.

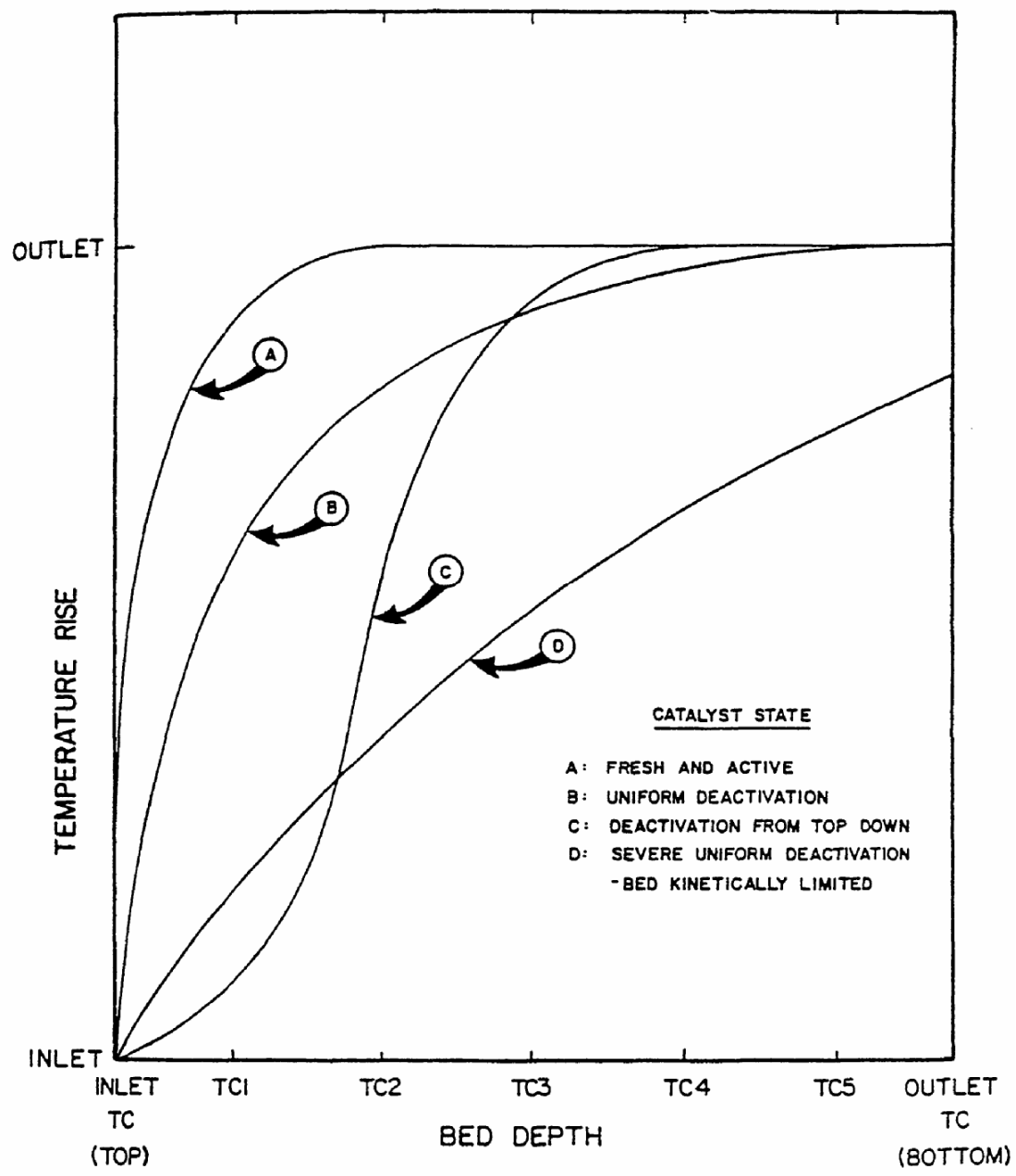


Figure 20.12: Converter Temperature Profiles Typical Of Various States Of Catalyst Deactivation

20.2.5.5 Design Criteria

The usual converter vessel currently used in sulphur plants is essentially a horizontal cylinder with gas inlet at the top and outlet at the bottom. The catalyst is supported as a 3 to 4 foot thick rectangular bed roughly centered on the cylinder axis. There may be a separate vessel for each converter, or the cylinder may be sectioned with longitudinal or radial inner walls to accommodate more than one converter in a single vessel. Another common arrangement is to stand a cylinder, sectioned with radial walls, on end so that the converters are stacked one over the other. In this case, the gas inlets and outlets are on the sides with gas flow down through 3 to 5 foot deep cylindrical catalyst beds. This arrangement saves ground space but does present inconveniences for catalyst placement.

All of the arrangements are characterized by a very low bed depth to width ratio. This results in very poor gas distribution through the bed and this significantly reduces the effectiveness of the catalyst within the central portion of the bed is effectively used.

The reason for choosing this "pancake" design was because of its low pressure drop. It seems that designers were unable, or unwilling, to give up the energy required to get good gas distribution. On the other hand, even now, the designing of converters to achieve the required distribution is very much an art. Nevertheless, good gas distribution within the converter is essential to obtain optimal use of the catalyst. As more precise data on the catalytic kinetics of the main and side reactions become available, the converters may be sized with more precision. The current tendency is to design with lower space velocities of the 700 to 1,100 hr^{-1} range based on total process gas at standard conditions and for most cases this appears to be adequate. However, significantly lower space velocities in downstream converters would be appropriate to offset the effect of the large reductions in reactant concentrations. The inlet reactant concentrations at first converters are 5 to 25 times the concentrations at downstream converters. The present practice of sizing all converters the same results in at most a 30 percent reduction of the actual space velocity due to lower operating temperatures.

The number stages required is of course a function of both the % H_2S in the acid gas and conversion required. Two to four are common. Data for actual plant performance is shown in Figure 20.13.

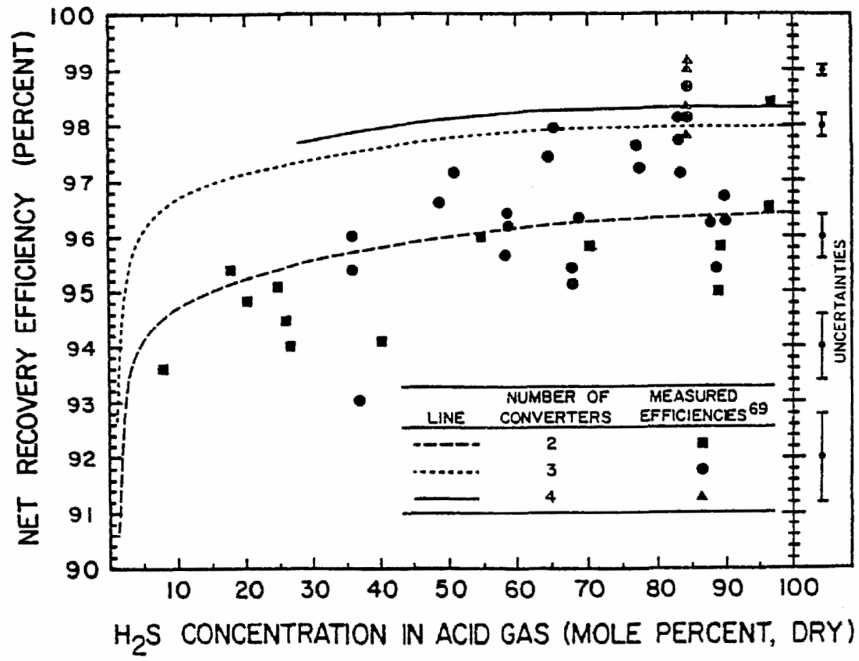


Figure 20.13: Practicable Capability the Modified - Claus Process as a Function of Acid Gas Composition

20.2.6 Sulphur Condensers

Sulphur condensers are built either as single-pass or multi-pass units (up to four tube passes in a single shell). In either case, they are generally designed with a minimum tube size of 1 inch. Normally 12 BWG steel tubes are used and installed with a 1/2 to 3/4-in, ligament between the outer walls of the tubes. Welded (ERW) tubes are generally used.

Sulphur condensers may be designed in accordance with ASME Section I or Section VIII Codes. In either case, tube-sheet design need only meet ASME rules of design to provide satisfactory service. Sulphur condenser tube attachments are typically rolled, seal welded, and rerolled.

Design mass velocity varies from 3.0 to 8.0 lb/sec/ft^2. Care must be taken to keep mass velocity sufficiently high during turndown conditions to prevent "sulphur misting". If "sulphur misting" occurs, liquid sulphur will not separate adequately from the process gas in the condenser separator chambers.

The liquid sulphur drain leg and nozzle should be steam-jacketed to prevent plugging by solid sulphur, and the drains should be installed for easy access for cleaning and disassembly. If possible, the sulphur condenser should be the low point in the Claus unit to allow any condensed sulphur to freely drain into the condenser and flow out through the drain legs.

Inspection ports and manways at each end of the condenser should be bolted to allow access to tubes for cleaning and inspection.

Vapour-liquid separator chambers are used to separate liquid sulphur from uncondensed process gas. These are made as either integral parts of the condenser or as separate vessels.

Most sulphur condensers are designed to generate steam on the shell side or used to preheat boiler feed water while the gases are cooled and sulphur is condensed on the tube side. Design steam pressure is usually 20 to 80 psig.

20.2.7 Sulphur Fog and Mist

Liquid fog or mist formation in sulphur condensers is a pernicious problem. It can occur and persist for long periods of time without the plant operator being aware of its existence.

The terms "fog" and "mist" are usually used interchangeably when referring to sulphur condensers. However, in the present context, a very subjective and semi-quantitative distinction is made. Droplets that are sufficiently small so as to behave as gas, and therefore not captured by normal coalescing or impingement devices, are considered as fog. This kind of behaviour has been confirmed with liquid $S0_3$ mist passing through filters and gas washing bottles containing H_2O. Mist droplets, on the other hand, are sufficiently large so as to be captured by mesh pads and packed coalescers.

Mist droplets may be formed by the growth of fog particles, or created by entrainment of liquid droplets into the gas stream from tube walls because of excessive gas speeds.

When fog or mist formation occurs unchecked in final condensers, it reduced plant recovery and contributes directly to the emission rate. Fog or mist from upstream condensers, if vaporized by the reheater or heat of reaction within the converter itself, affects the operation of the following converter in two ways.

Firstly, it elevates the sulphur dewpoint so that a higher operating temperature is required to maintain the same dewpoint margin. Secondly, the increased sulphur vapour shifts the H_2S-$S0_2$ equilibrium reaction to the left. Both these effects reduce conversion.

If the droplets are not vaporized, they deposit on the catalyst and gradually deactivate the bed. Measured evaporation rates of liquid sulphur droplets indicate that the droplets are not likely vaporized by the time they reach the converter.

In a following section, the effect of sulphur recovery efficiency on conversion is illustrated. Because the conversion loss due to increased sulphur vapour partial pressure is very small compared to the loss to be expected by deactivation of the bed, non-vaporization of the droplets is much more serious than when they are vaporized.

The present consensus is that the best way to handle sulphur fog is to prevent it. Fog occurs when the gas stream cooling curve crosses the dewpoint curve, resulting in super saturation of the gas. Condensation then takes place by the formation of small droplets within the gas stream rather than as a film on the cooler tube surface. These conditions arise when the heat transfer rate from the gas stream is too high or when temperature gradients across process streams are very large. Fortunately, the conditions are sufficiently well known that it is possible to design condensers to avoid the problem.

Although sulphur fog will form without condensation nuclei present, the presence of such nuclei dramatically enhances fog formation. Thus, keeping particulate matter, such as catalyst fines or metal corrosion products, out of the gas stream is an important step in preventing fog formation.

20.2.8 Coalescers and Mist Eliminators

The importance of using of coalescers and mist elimination devices in sulphur plants has been overlooked. This is clearly demonstrated by the number of plants that were built and operated without them. However, as air pollution regulations became more restrictive, operations began to realize the value of such devices as they attempted to fore higher recoveries from their plants. Still, some operators saw no apparent benefits from their coalescers and further, did not appreciate them as insurance devices. Some observed dramatic results, such as: 2 percent of the total plant production coming off the coalescers; a 0.5 percent loss in recovery due to corroded mesh pads in

the final condenser-coalescer; and converters falling below the sulphur dewpoint because of sulphur mist carry over from the preceding condenser.

To achieve maximum sulphur plant performance, it is essential to have mist extractors in all condensers and a separate coalescer following the final condenser of each train. It must be recognized that these devices are like insurance policies - their value is not fully appreciated until they come into effect.

Packed coalescers, possibly backed up with mesh pads, are preferred over open vessels with only mesh pads at the exit. For mist extractors in condensers, the horizontal arrangement is recommended over the sloped arrangement. For both, the linear gas speed is a critical design parameter.

A priori estimates of liquid sulphur entrainment are very difficult to make with a high level of confidence. Since adequate methods for direct measurement are not available, the uncertainties of a postiori estimates are usually large. Entrainment is usually defined for measurement purposes as the difference between the recovery efficiency measured directly on the stack gas and the recovery efficiency determined from the measured conversion efficiency and the calculated vapour loss at the final condenser. This difference of two large numbers method leads to the large uncertainty - often the uncertainty is larger than the determined entrainment.

Confidence in the result can be increased, however, by simultaneously determining the conversion and recovery efficiencies by as many independent methods as possible. A critical step in entrainment investigation is to examine and trace all piping entering the incinerator to ensure that an extraneous source of sulphur does not invalidate the measurements.

It has been stated that for a well designed sulphur plant the liquid sulphur mist entrainment will be no more than 2 to 3 pounds per 100 pound-moles of sulphur-free gas. This corresponds to maximum efficiency losses of 0.3 and 1.1 percent for pure and 10 percent H_2S feed streams, respectively. While measurements at a large number of plants indicated losses by this mechanism up to 2 percent, the data also show that in general these losses are preventable by careful operation, use of inter-stage mist eliminators and well designed coalescers. A number of previously diagnosed entrainment problems were shown to be sulphur (H_2S) in the incinerator fuel gas or "garbage" streams being disposed of in the incinerator.

Losses due to liquid sulphur entrainment have been reduced by careful plant design and operations to less than 1/2 pound of sulphur per 100 pound-moles of sulphur-free gas. For 10 percent and pure H_2S feed streams, this level translates into maximum efficiency losses of 0.2 and 0.05 percent, respectively.

20.2.9 Incinerator and Stack

Experience has shown that a steel high temperature refractory lined stack is by far the best from a corrosion point of view. It is designed to keep the steel below softening point but above dewpoint of gases. A typical design is as shown in shown in Figure 20.14.

Note: The Government of Alberta has some very special regulations in regard to stacks.

Generally they are:

Minimum exit temperature 1,000ºF.

Stack height is based on making sure ground concentrations of SO_2 are not above 0.17 ppm for 1 hour.

These regulations and the sizing of stacks is reviewed in Chapter 22 "Environmental Considerations". Other details on the stack are given below.

All sulphur compounds can be oxidized to SO_2 both thermally and catalytically. Thermal oxidation normally occurs at temperatures between 1,000ºF and 1,500ºF in the presence of excess oxygen. The penalty for using insufficient oxygen and/or lower temperatures is that some H_2S may not be converted to SO_2; thus H_2S is discharged to the atmosphere. Using too much oxygen consumes costly fuel.

Claus unit tail gas will not support combustion because there are too few combustibles present (< 3%). Thus, oxidation of all sulphur components in tail gas to SO_2 requires use of external fuel and air.

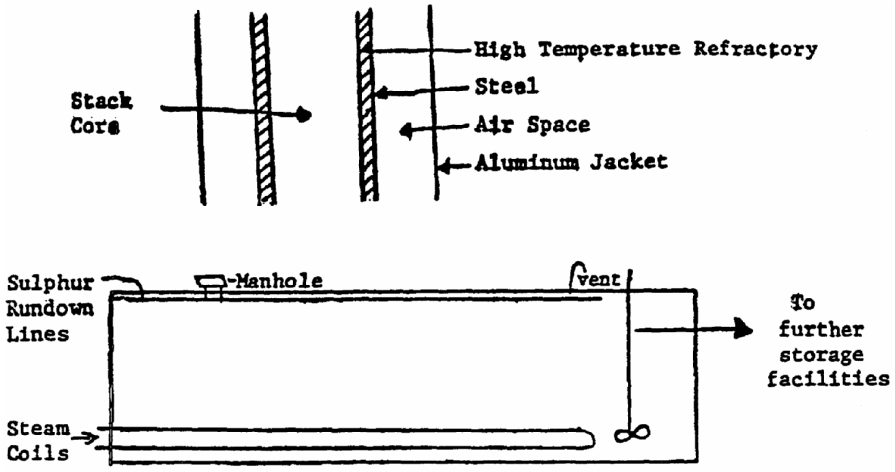


Figure 20.14: Incinerator and Stack

Catalytic incineration, a recent development, works when tail gas is heated to 700ºF-900ºF. Heated gas and heated air in contact with a fixed-bed catalytic reactor promote oxidation of sulphur components to SO_2.

Air is fed by natural draft for most thermal incinerators, but for those operating with waste heat recovery sections, a blower is used. Usually at least 25% excess air is used for proper tail gas incineration.

Incinerator and stack are often combined into a single vessel called an incinerator-stack. In this convenient design, the incinerator is usually an enlarged base and the stack of a small diameter extending upward from the incinerator. A horizontal burner normally mounts into the incinerator side. Tail gas can be fed either through the burner or into the incinerator adjacent to the burner via another nozzle.

Normally the incinerator is sized for at least 0.5 - 1.5 seconds residence time.

Most incinerator-stacks are constructed with a steel shell lined with heat-resistant refractory. Such stacks can be free-standing, or guy-supported or derrick-supported structures. Selection of support type is dictated by stack height, wind load and/or seismic criteria for the site. A guy-supported stack is most economical; however, a larger plot space is needed for guy anchors and cables. A free standing stack is most commonly used for heights not exceeding 250 ft.

Incinerators are generally designed for a maximum operating temperature of 2,000ºF. Thus refractories rated as low as 2,200ºF have been successfully used in incinerator-stack installations. Normally, one layer of refractory is sufficient, and is usually an insulating castable rated at service conditions of 2,200ºF or above.

Vertical incinerators installed at the base of the stack require proper protection for the incinerator steel floor. The preferred protection is circulation of ambient air below the steel floor. For best

results, the steel should be kept cooler than 650ºF-700ºF, and hotter than 300ºF, water can condense against the shell, absorb SO_2 and form a very strong, highly corrosive acid. Accomplishing these temperature goals requires careful design of both inside refractory lining and outside insulation. Normally, refractory thickness to protect the incinerator shell (either vertical or horizontal type) is 2 to 4 inches. The incinerator floor on vertical units normally employs 4 to 6 inches of refractory.

The incinerator-stack should be insulated on the outside to prevent low shell temperature. This insulation can be an air gap (usually 3 to 4 inches deep) with either a stainless steel or aluminum sheathing around the incinerator-stack, or a < 1.0 inch thick layer of blanket-type insulation covered with aluminum or stainless steel. Over-insulation can cause steel shell overheating.

Most government air control units (federal, or provincial) require that emissions be monitored on a periodic basis. Thus the incinerator-stack is usually equipped with platforms, sample nozzles and utilities located at the proper level. This equipment permits periodic measurement of stack flow rate and stack gas composition for use in determining sulphur emission rates.

Where continuous monitoring is required, SO_2 stack gas analyzers are installed on the stack. These instruments are often coupled with flow meters to develop a continuous record of flow and SO_2 content.

Typical stack velocities for design are 40 to 100 ft/sec, with allowable pressure drop generally determining actual design flow rate.

20.2.10 Other Design Considerations

- Make sure air blower is sized for air requirements with some hydrocarbons present; an allowance of 2% hydrocarbons in the acid gas is a good basis for sizing the air blower.
- Keep the acid gas temperature to the sulphur plant as low as possible to knock out all water vapour in the acid gas feed stream as H_2O is a product of the reaction. The feed gas can be heated after minimum temperature has been reached and water removed.
- Do not use high alumina cement for liquid sulphur storage tanks. It loses all strength under moist warm conditions more quickly than ordinary cement.
- Make sure all sulphur drains can be properly rodded.
- Make sure all drains are visual.

20.2.11 OPERATIONS

- Keep the ratio of H_2S to O_2 as close to 2 to 1 as possible.

  This can be done by measuring the H_2S and SO_2 in the tail gas and re-adjusting the air so that this is always 2 to 1. This is now being done by a chromatograph. Figure 20.15 shows the effect of being off ratio.

- The sulphur condenser outlet temperature should be as close to 270ºF - 280ºF as possible.

  With very much lower temperatures sulphur begins to freeze and with very much higher temperature sulphur vapour is left in the gas and also it becomes very viscous between 325ºF - 380ºF.

- Watch the temperature rise across each converter closely to see the amount of reaction taking place in the converters.

  A sudden decrease should be investigated to see that cause of same.

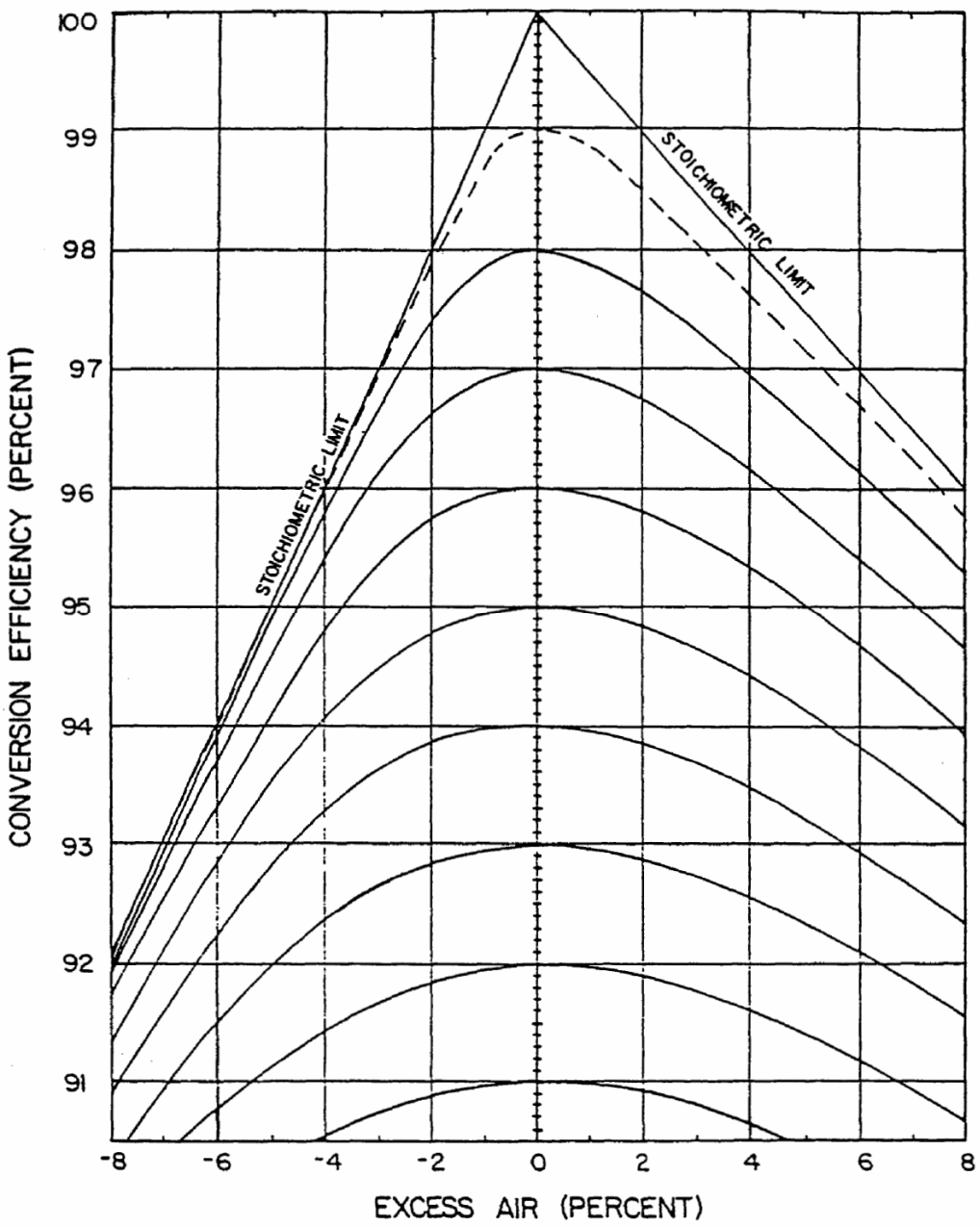


Figure 20.15: Efficiency Loss as a Function of Excess Reaction Air (Modified-Claus Process)

20.2.12 Plant Stability

It has been a consistent general observation that the highest recovery efficiency is achieved by a plant during stabilized, steady-state operations. Any change in process conditions temporarily

decreases conversion efficiency. The change is judged good or bad by noting whether the restabilized conversion is higher or lower than before the change.

For any given set of operating conditions there is an upper limit‡ imposed on the efficiency by thermodynamic equilibria and possibly kinetic limitations. A plant that operates under truly stabilized steady-state conditions will achieve this "upper limit" efficiency. But, any change in operations disturbs the system stability and the immediate response for sulphur plants is at least a momentary decrease in the efficiency. If no further changes are made, stability will eventually be re-established and the efficiency will level off at the new upper limit efficiency determined by the resulting operating conditions.

These effects are illustrated in Figure 20.16 which is a continuous record of the recovery efficiency of a three-stage plant. The record was obtained during field trials of a prototype sulphur plant control system/optimizer. Control of the air to acid gas ratio was switched to the prototype control system at the beginning of the time interval covered. All other sulphur plant operations were normal; changes were not initiated by the operator. The net increase of 0.45 percent in the efficiency was due to the better air to acid gas ratio control.

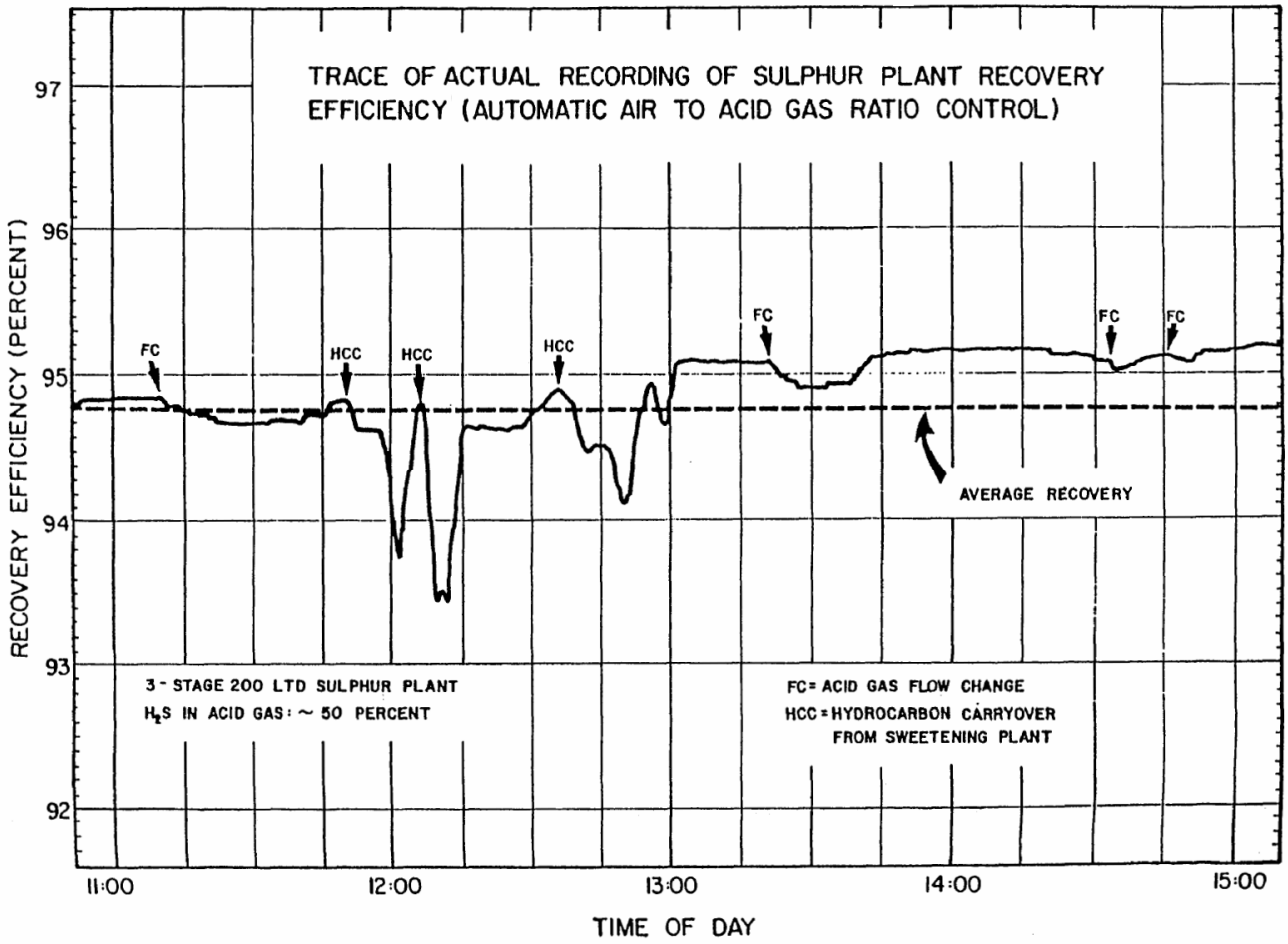


Figure 20.16: Trace of Actual Recording of Sulphur Plant Recovery Efficiency (Automatic Air to Acid Gas Ratio Control)

‡ The upper limit conversion efficiency discussed here is not an absolute.

The effects of plant load changes and slugs of hydrocarbon from the gas sweetening process on efficiency are clearly evident in Figure 20.16. It is a variable determined by the plant operations conditions - load (space velocity), temperatures, pressures, reactant concentrations, catalyst condition (exist at a given instant in time). The limit will approach or equal the optimal thermodynamic or realizable efficiency only if the stabilized steady-state operating conditions coincide with the most ideal conditions.

The efficiency decreases are manifestations of the destabilizing effects both events have on the air to acid gas ratio and the converter temperatures and flow distributions. For the load changes the air flow must be adjusted to match the change in acid gas throughput. The hydrocarbons consume some of the O_2 resulting in deficient air for the Claus reactions. Both events cause changes in temperatures and space velocities in the converters.

The time required for restabilization of the plant depends upon the type and magnitude of the change, size of the reaction vessels and the effectiveness of the plant control system.

The acid gas flow rate to any plant is never absolutely steady. This makes it difficult to maintain truly stable operating conditions, particularly since the air flow must follow the ups and downs of the acid gas flow. Following the installation of new ratio control systems, operators have reported larger increases in conversion than can be accounted for by the improved ratio control. This is likely a benefit derived from the more stable converter and condenser conditions resulting from the better air to acid gas ratio control. A further benefit of the stability resulting from good ratio control is savings in incinerator fuel since lower levels of excess O_2 are required in the stack to assure complete oxidation of the H_2S.

Frequent and/or large load changes reduce plant performance, although it is difficult to quantify. Because of this it has been common practice in the past to operate multi-train plants with all but one train controlled at constant load. The remaining train is then left to take the inevitable large load swings.

Early attempts at establishing a correlation between efficiency and plant load failed because the efficiency is affected more by other factors associated with the load change, such as changes in air to acid gas ratio and changes in vessel temperatures and pressures, than by the load change per se. One series of tests at a two train plant showed that, given sufficient time to re-establish ratio control and equilibrium in the vessels, the difference between plant conversions at 60 and 110 percent of design load was insignificant. For well designed plants, this is the expected result. For plant where the efficiency is kinetically limited by undersized vessels, the efficiency would increase as the plant load is decreased.

The necessity for operating converter outlets above rather than at the sulphur dewpoint temperatures results from a combination of design and control limitations. Uncertainties on temperature measurements, uncertainties on sulphur dewpoint temperatures, lack of knowledge of instantaneous conversion efficiencies, frequency of plant load changes and process upsets are all considered when selecting the operational dewpoint margin and correspond to a 15ºF dewpoint margin.

Sulphur fog or mist entrainment is determined indirectly as the small difference between two large numbers, the conversion and recovery efficiencies. Therefore, the uncertainty on the entrainment is large. As noted before, measurements have shown that the entrainment in well designed and controlled plants can be reduced to levels well within its measurement uncertainty. Therefore, the realizable minimal losses for this factor in Table 20.3 are the measurement uncertainties estimated previously.

Efficiency losses due to air to acid gas ratio control are independent of feed composition, but are dependent upon the conversion efficiency. They are determined primarily by the effectiveness of

the control system although they can be dramatically increased for very short periods by upsets in the sweetening plant, frequent sudden load changes and perhaps limitations of the air supply system. For highly efficient, well controlled plants this loss has been shown to be reducible to less than 0.2 percent on the short term. For Table 20.3, it was assumed that the most efficient control system available is used to control the air to acid gas ratio to within ±1 percent excess air based on optimal conversion (Figure 20.15).

The "practicable recovery efficiency" in Table 20.3 is that which can be achieved in the longer term taking into consideration the fact that various operational realities influence the just previously discussed losses:

- Transitory upsets in the sweetening process can reduce the efficiency by causing momentary deficient air operation. The hydrocarbon carryover episodes noted in Figure 20.16 reduced the average efficiency for the recorded interval by about 0.2 percent.
- Deterioration of the catalyst.
- Corroded mesh pads in the final condenser/coalescer which can lead to large efficiency losses from sulphur fog or mist entrainment.
- Primary sensing devices (thermocouples, pressure gauges, flow measuring elements) drift off calibration as a result of corrosion or wear.
- Deterioration of the plant equipment, such as burner tips, control valves, heat exchanger fouling, line plugging.
- Operational problems in the sweetening process such as poor pressure control, entrainment of liquid in acid gas stream or malfunctions in condensers or coolers.
- Variations in ambient temperature and humidity as they affect air blower efficiency and the thermodynamic conversion efficiency.

Precise quantifications of the above factors are not possible, particularly since they are so highly variable. Moreover, they in general do not all occur simultaneously. Judging from the review of long term performances of well run plants where reliable and consistent data were available, it is not unreasonable to expect the efficiency loss from the above factors for rich feed plants to be equal to the sum of the realizable minimum losses discussed earlier.

By virtue of the acid gas bypass and lower reaction furnace temperatures, lean feed plants are much more susceptible to catalyst deactivation. Since catalyst deactivation can seriously limit recovery efficiency, the loss from the above group is estimated as double the sum of the minimum realizable losses.

Table 20.3: Practicable Sulphur Recoveries by the Conventional Modified-Claus Process

Item	Rich Feed (85% H_2S)		Lean Feed (18% H_2S)
	4 Converters (% of Input)	3 Converters (% of Input)	3 Converters (% of Input)
Ideal Recovery	**99.50**	**99.19**	**99.89**
Minimum Realizable Losses due to:			
Operation above dewpoint.	-0.12	-0.18	-0.16
Sulphur fog/mist.	-0.08	-0.15	-0.18
Air to acid gas ratio fluctuations.	-0.39	-0.29	-0.23
Realizable Recovery	**98.91**	**98.57**	**98.32**

Estimated loss due to other operational factors.	-0.60	-0.60	-1.20
Practicable Recovery	**98.31**	**97.97**	**97.12**

20.2.13 Operating Problems

It is very easy to operate a sulphur plant. Problems that arise include:

1. Foaming in the sweetening plant. This causes surge of feed gas and plant is difficult to control. Also hydrocarbons and sweetening solutions can be carried over.
2. Plugging of sulphur rundown lines; too low steam pressure or shutting off steam causes this problem. The lines are usually rodded to clear them.
3. Sulfidization at temperatures over 650ºF the reactants H_2S and SO_2 often act on steel and form pyrites. It is best to line all pipe vessels that may be exposed to temperatures above 650ºF with refractory to get over this problem.
4. Leaking condensers are often where the tubes leak around the tube sheet due to unequal expansion. Some people seal weld these joints but not everyone.

Note: A good review of sulphur plant operation is given by Reference "17".

20.2.14 Other Reactions Occurring in Process

1. All hydrocarbons generally are burned to CO_2 + $2H_20$:

Reaction 20-5

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

but unfortunately sometimes, particularly with heavier hydrocarbons, carbon is formed. This carbon, of course, plugs up the catalyst pores.

2. Mercaptans generally form elemental sulphur as follows:

Reaction 20-6

$$CH_4S + 2O_2 \rightarrow CO_2 + 2H_2O + S$$

The breakdown of COS and CO_2 has been given previously. However COS can be formed in the reaction furnaces. i.e.

Reaction 20-7

$$CO_2 + H_2S \rightarrow COS + H_2O$$

The sulphur species shown in the above equation have been shown as single molecules of S. Often, S_2, S_4, S_6 are formed.

20.2.15 Emergencies

20.2.15.1 Hydrocarbon Carryover

Slugs of hydrocarbon can, and often do, carry into the amine system. These can be removed in the amine regenerator and end up in the acid gas. If the hydrocarbon is methane, the reaction occurring in the reaction furnace is:

Reaction 20-8

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

If the hydrocarbon is butane the reaction is:

Reaction 20-9

$$C_4H_{10} + \frac{5}{2}O_2 \rightarrow 4CO_2 + 5H_2O$$

The butane requires over three times the amount of oxygen that methane does. As the reaction furnace reaction occurs with insufficient air we are likely to get the following reaction instead:

Reaction 20-10

$$C_4H_{10} + \frac{5}{2}O_2 \rightarrow 4C + 5H_2O$$

This reaction requires less than half as much oxygen but results in the formation of coke. Coke plugs off the catalyst decreasing efficiency and increasing pressure drops.

When a hydrocarbon slug hits the reaction furnace, large quantities of air are used to burn the hydrocarbons. The result is that less oxygen is available to burn the H_2S to SO_2. This H_2S gas goes through the unit and burns in the incinerator resulting in a high stack temperature. The stack temperature can be decreased by increasing the air to the reaction furnace and getting the combustion in the reaction furnace where it belongs. Obviously, the source of hydrocarbon should be cut out immediately.

20.2.15.2 Flame Failure

Flame failure in the reaction furnace can result from a wide swing in the acid gas flow or from a slug of water from the knockout drum. If the fire is lost, the oxygen is not burned in the reaction furnace and carries through to the converter bed. The result is a catalyst fire. Bed damage can occur within minutes if it is not brought under control.

To control the catalyst fire, cut out the air immediately, then the acid gas. Purge the reaction furnace, then relight the fire and bring the unit back on line. Also often steam spargers are put in the converters for snuffling catalyst fires.

20.3 Storing and Loading of Sulphur

20.3.1 Introduction

Sulphur is drained off the various condensers, converters, etc. into a covered concrete pit that is filled with steam coils at the bottom of it. Figure 20.14 is a sketch of the facility. Also insulated surface tanks are sometimes added as supplemental storage. It should be noted that there are special problems in storing liquids. Liquid sulphur absorbs H_2S.

Generally at least 3 days production is allowed for in sizing this pit and often more if most of the sales are to be liquid.

If liquid is the primary method of moving sulphur from the plant, then sometimes insulated heating surface tanks are installed to supplement the concrete pits.

In most Alberta plants, sulphur was originally stored as a solid in large vats approximately 30 feet high. It is generally poured on these vats in layers about 6" thick to allow the sulphur to cool so that local hot spots do not occur. The sulphur was then broken up with bulldozers and carried by conveyors into boxcars or gondolas. This type of process is causing a great deal of dusting and

results in some contamination of the surrounding countryside. For example, soil pH of 2.0 has been measured in the vicinity of sulphur storage piles. Presently the block storage or liquid production is made into prills or slates.

The Canadian Petroleum Association has developed a sulphur Handling Guide that covers most of the storing and handling problems and methods except for prilling which is discussed further on.

Liquid sulphur is pumped and handled in more or less the same manner as any other liquid except that it must be kept heated to prevent it from freezing. This occurs at approximately 115ºC (240ºF), so that it is normally maintained in the range of 130ºC to 150ºC (265ºF to 300ºF). This usually involves the use of steam coils in tanks, steam jacketed, insulated piping, etc., although other heating media are also used. Liquid sulphur is shipped in rail tank cars and tank trucks. These are insulated and equipped with steam heating coils so that the sulphur may be thawed if necessary at it destination, prior to unloading.

Most of the large plants (with design sulphur production capacities or over 900 t/d) have solid sulphur forming and loading facilities. Smaller plants normally ship only liquid. Any excess of sulphur production over shipments, after liquid storage and any formed sulphur storage are full, is sent to sulphur block storage.

20.3.2 Sulphur Block Storage

Solid sulphur is stored in large blocks of varying areal extent and approximately 9 to 10 m high. These are formed by pouring liquid sulphur in thin layers and allowing each layer to solidify before pouring additional liquid in that area. This avoids the encapsulation of deep pools of liquid in the solid block. Normally several pouring spouts are used on a rotating basis over a sufficient areal extent to accomplish this. Aluminum or plywood slip forms are generally used around the perimeter of the pouring area to hold the liquid until it solidifies.

The general practice has been to use the initial sulphur poured as the base pad for the block. The sulphur is reclaimed from the block to the level of 0.3 m to 1 m from the ground, leaving a pad to be reused for the formation of new blocks. The disadvantage of this practice is that when the block storage area is ultimately to be abandoned and the block pad to be removed, a mixture of sulphur and soil must be reclaimed. The amount of this mixture and the difficulty of separation will depend on the soil type, and the degree of site preparation that was carried out prior to pouring the original sulphur. This latter factor varies widely. In some cases there has been minimal preparation; in others the removal of top soil and organic debris, and grading and levelling has been carried. Some gravel pads have been laid. A limited number of concrete and asphalt pads have been installed, particularly at some of the new plants.

Prior to about 1970 all solid sulphur was shipped as "bulk" sulphur. This was obtained directly from the block by breaking up the sulphur and crushing the larger lumps to a maximum size of about 20 mm. Hence, little or no melting was practiced. This method of shipment, however, involved sulphur dust problems, particularly in handling overseas export shipments through the port of Vancouver, which became increasingly serious as tonnages increased. This led to the installation of sulphur forming facilities at the larger plants to handle these export volumes. As a result, only a small amount of solid sulphur is still shipped as "bulk" sulphur to Canadian and U.S. markets only. Since the early seventies, therefore, the solid sulphur reclaimed from the block is normally melted to provide liquid feed to sulphur forming facilities and/or liquid shipments.

The sulphur melters currently in use are of two types, "static" and "insitu". Static melters are usually located immediately adjacent to the block. The solid sulphur is mechanically broken from the block, and is transported and fed to the melter, typically by mobile equipment such as a front-end loader. The tube type melter in common use consists of a number of vertical steam jacketed

tubes surmounted by a loading hopper. The sulphur melts on the walls of the tubes and the liquid runs off to a surge tank, from which it is pumped. These facilities are normally designed for easy relocation to other areas of the block, but they are static while in service.

The insitu-melter is designed to actually bring a steam heated steel surface in contact with the sulphur in the block. The liquid runs down the surface and is collected. The insitu-melter is constantly moving slowly against the surface of the block as it is melted. This type of melting has the advantage of not requiring mechanical break-up of the block, thereby essentially eliminating dust formation during reclamation.

20.3.3 Sulphur Forming

The two sulphur forming processes which account for the majority of tonnage formed in Alberta are the slating process, and the 'Ciech'air prilling process which was developed in Poland. There are currently an equal number of slating and air prilling facilities in operation in the province, however, all of the air prilling facilities have been installed since early in 1980. There are also a few smaller facilities utilizing wet prilling and granulation processes.

The slating process introduced in the early seventies, was the first sulphur forming process to be used commercially in the province. It is illustrated diagramatically in Figure 20.17. The liquid sulphur is pumped to the top of the precooler, where it is cooled by direct air contact to about 1250ºC (260ºF). It flows by gravity to distributors located at the inlet end of each slating belt. These are flat moving belts made of rubber of stainless steel having in the order of 30 to 40 meter effective length. A typical installation will consist of 8 to 10 or more belts operating in parallel. The liquid sulphur is distributed over the width of the belt and is cooled by air from above and water from below. By the time it reaches the end of the belt, it has solidified into a single sheet. This sheet breaks off the belt into individual slates about 10 to 12 mm thick and having an irregular shape perhaps 100 mm across. It is then conveyed to storage, which is usually designed for rapid reclaim and loadout of rail cars. Storage of formed sulphur is limited, typically to two or three train loads or less.

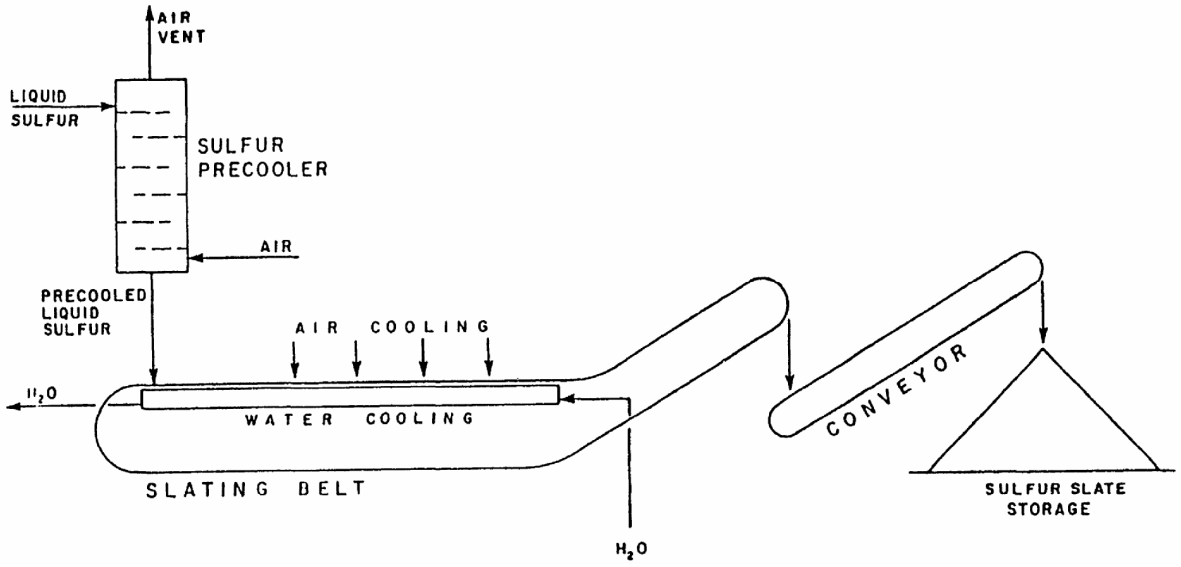


Figure 20.17: Typical Sulphur Slating Process

Sulphur slates, although an improvement over bulk sulphur, are still subject to attrition and fines production during handling, due to the less than optimum slate geometry. This has led to the introduction of prilling and granulation processes which produce small spherical shapes.

The Ciech sulphur air prilling process is illustrated diagrammatically in Figure 20.18. Liquid sulphur is pumped to the top of the prilling tower and is sprayed through 2mm holes to form liquid droplets. The sulphur droplets are cooled by a countercurrent stream of air during a drop of form 30 to 60 m to the collection grid. This results in a generally spherical product 1 to 6 mm in diameter, which is conveyed to storage facilities similar to that of slate storage.

Wet prilling process use water as the cooling medium instead of air. Granulation processes rotate a bed of fine solid spherical particles past a fine spray of liquid sulphur, in a cooling air stream. This produces a spherical granule in the required size range, as successive layers of liquid adhere to and solidify around the nucleus.

Another process is Rotoforming as shown in Figure 20.19. Liquid sulphur is extruded in small drops through a Rotoforming head on to a moving cooled stainless steel cooling belt. The drops as they move along the belt are cooled and solidified by a water spray on the underside of the belt. The solid particles drop off the belt (helped by a blade) on to a conveyor belt that carries them to a storage silo.

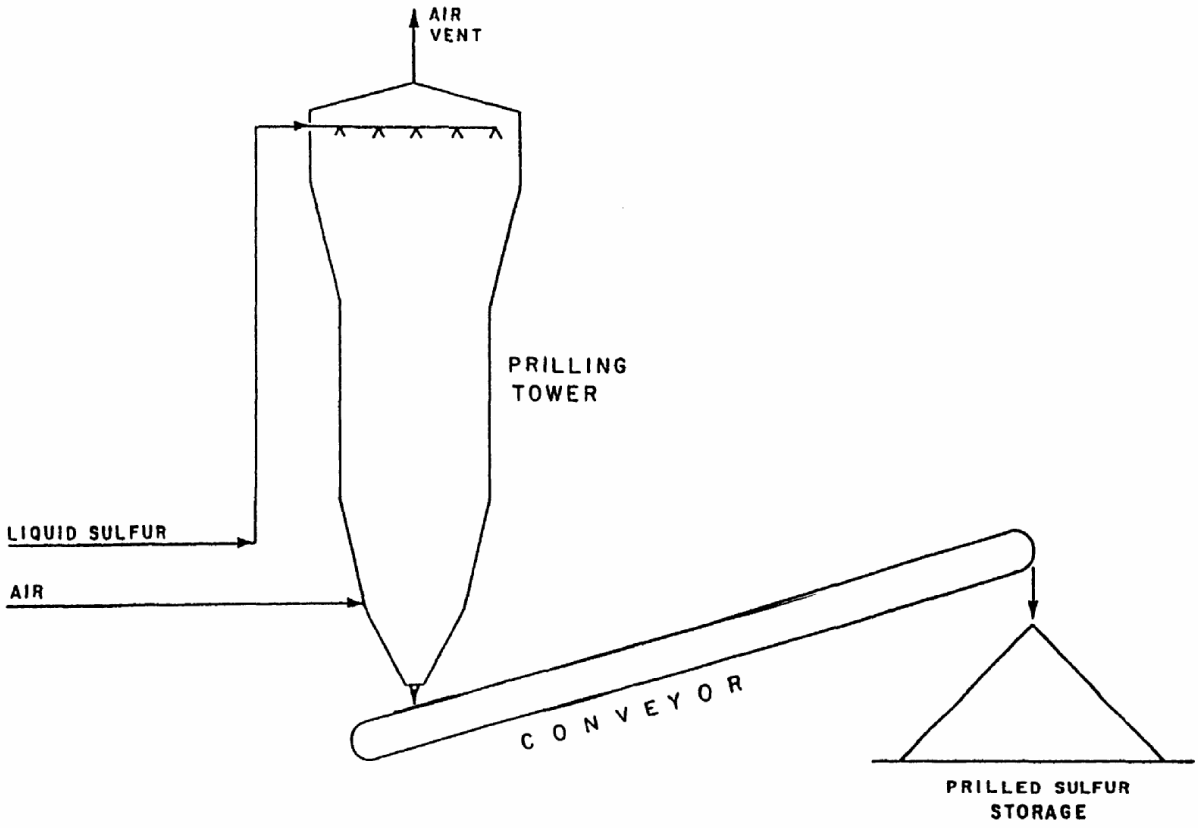


Figure 20.18: Sulphur Prilling Facility

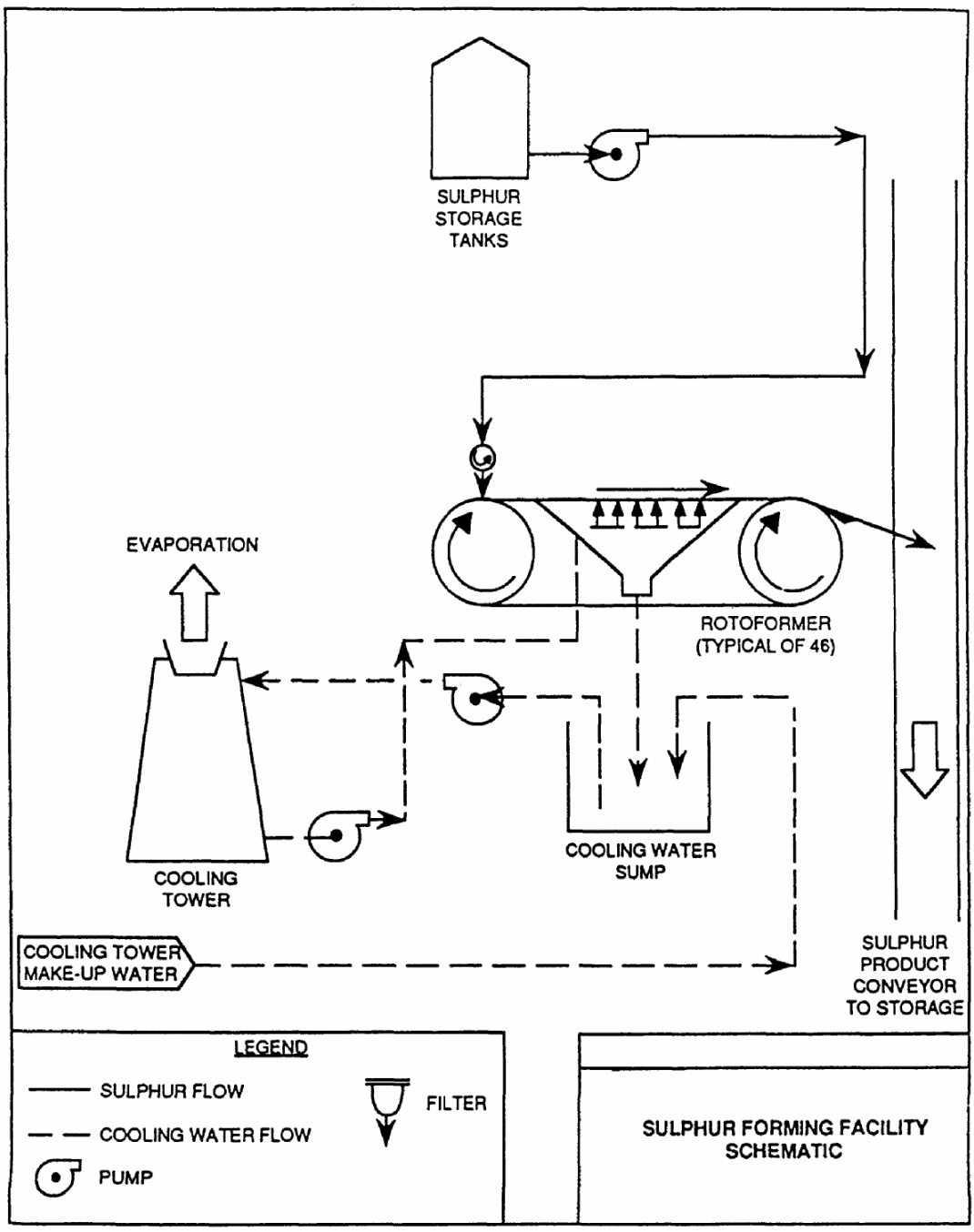


Figure 20.19: Sulphur Forming Facility Schematic

20.4 Sulphur Plant Tail Gas Clean-Up Processes

20.4.1 Introduction

Because of the more stringent requirements of Pollution Control, requirements for tail gas clean-up process has been developed. For example, the present ERCB guidelines are shown in Figure 20.20.

The requirements as published by ERCB-AE-88AA19 is as follows:

New plants with sulphur inlets of 2,000 tons per day or more have had sulphur recovery requirements increased to 99.8% from 99%.

Those in the 50 to 2,000 t/d range must recover between 98.5% and 98.8%, up from the previous 96.2% to 98.8%.

New plants with inlets of 10 to 50 t/d must recover 96.2%, up from the previous range of 92% to 96%.

The ERCB has also decided to require sulphur recovery on small plants of one to 10 t/d. Previously, there were no recovery requirements on these plants. The new rules now call for a minimum 90% recovery level for plants between five and 10 t/d, and a 70% recovery for plants between one and five t/d.

The basis for this decision is given in the above referred to report and in more detail by Reference 19.

The technology in tail gas cleanup is very fluid as there is extensive work done on process for SO_2 removal from power plant stack gases as well as for sulphur plants. Generally sulphur plant owners favour the process in which elemental sulphur is recovered: The previously mentioned Reference 19 is a good review on the subject.

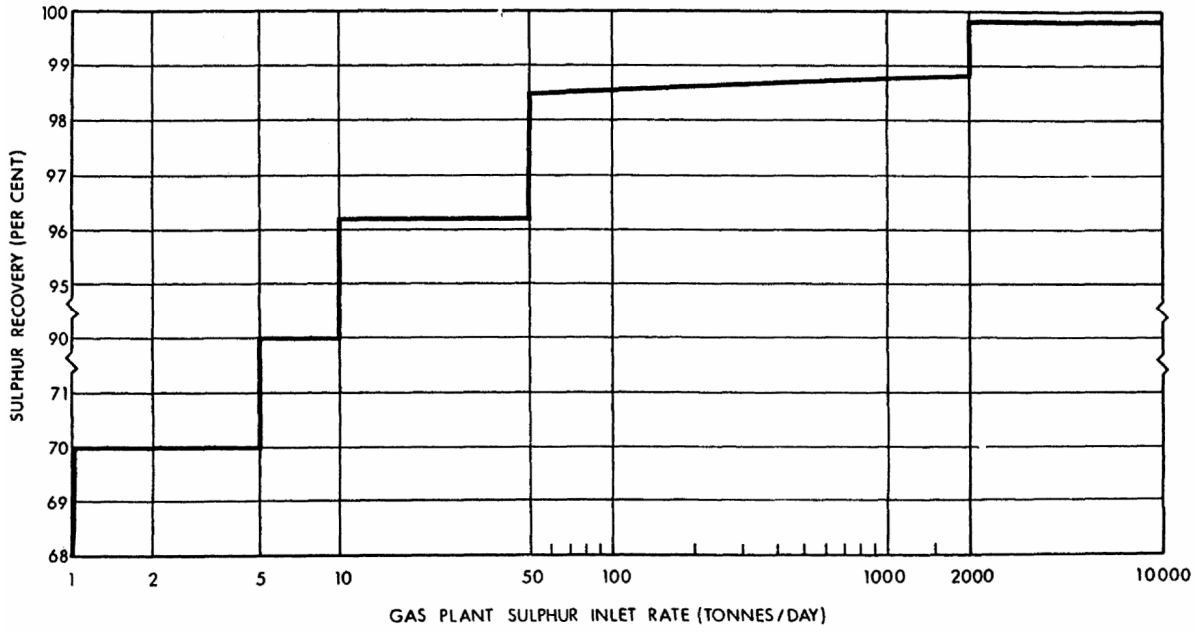


Figure 20.20: Sulphur Recovery Requirements for Alberta Sour Gas Plants - August 1988

20.4.2 Processes

Because the Claus process is unable to convert 100% of the H_2S in the acid gas to elemental sulphur and because of the concerns about emitting SO_2 to the atmosphere, various processes have been developed to treat Claus plant tail gas to remove residual sulphur compounds that would otherwise be emitted. Some of these processes continue the Claus reaction on catalyst beds operated at sub-dewpoint temperatures to enhance the conversion of H_2S and SO_2 to elemental sulphur. Another group of processes designed to treat Claus tail gas convert all sulphur compounds present to H_2S by hydrogenation and hydrolysis.

The H_2S is then absorbed out of the tail gas and either recycled to the front of the Claus plant or is reacted directly to form elemental sulphur. Still another group of processes oxidizes all sulphur compounds present in the Claus tail gas to SO_2 which is then removed from the tail gas and recycled to the Claus plant or is reacted chemically to form a by-product. It is possible with some of these tail gas cleanup processes to increase the overall sulphur recovery to essentially 100% on a normal operating basis. The following sulphur recovery processes can be considered:

Primary

- Modified-Claus
- COPE
- Linde Oxygen Enrichment
- MCRC
- Selectox/Recycle Selectox
- Stretford Sulphur Removal
- Lo-Cat
- Kellog Oxydesulphurization
- Stauffer Sulphur Recovery
- Integrated UCAP

Tail Gas Cleanup	
• Sulfreen • Amoco CBA • MCRC • Maxisulf	Continue Claus reaction at sub-dewpoint temperatures.
• IFP Clauspol 1500	Continues Claus reaction in liquid phase.

• SCOT • BSR/MDEA • BSR/Stretford • BSR/Unisulf • Sulften • BSR/Selectox • Oxysulfreen • ULTRA	Convert sulphur compounds in Claus tail gas to H_2S for selective absorption and recycle to Claus plant or for direct oxidation to sulphur or for partial oxidation to SO_2 and continuation of Claus reaction.
• UCAP • Wellman-Lord	Convert sulphur compounds in Claus tail gas to SO_2 for selective absorption and recycle to Claus plant.
• Citrate	Converts sulphur compounds in Claus tail gas to SO_2 for Claus reaction with H_2S.
• ATS	Converts sulphur compounds to SO_2 for reaction with ammonia.
• Pritchard Clean Air	Selective absorption of SO_2 and H_2S from Claus tail gas for extension of Claus reaction and liquid phase oxidation of H_2S to sulphur.

Not all of the above processes have been used commercially. As well, some of the above processes are in an advanced stage of development but not yet commercially available.

The commercially tried processes are shown in Figure 20.21. A good reference on these processes is given by Goar. The processes that have been used or pilot plant tested in Western Canada are the following:

- Sulfreen Process
- Amoco CBA
- IFP Process
- SCOT Process
- Beavon
- Pritchard Clean-Air
- Wellman-Lord Process
- Lo-Cat Process
- MCRC Process

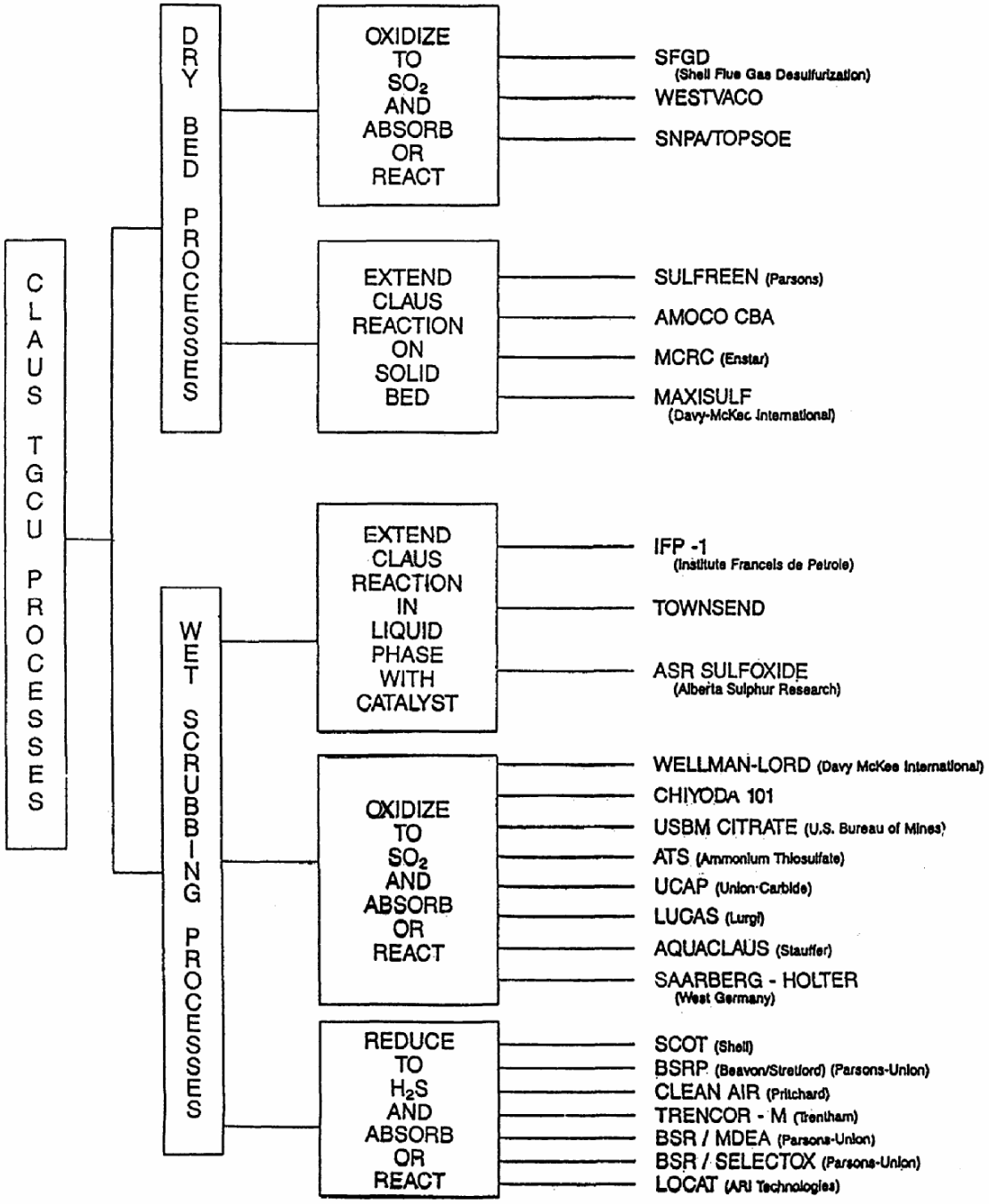


Figure 20.21: Tail Gas Clean-up Processes

20.4.2.1 The Sulfreen Process

The Sulfreen process converts H_2S and SO_2 contained in the tail gas to sulphur at low temperatures (260ºF - 300ºF) by extension of the classic Claus reaction:

Reaction 20-11

$$2H_2S + SO_2 \xrightarrow{catalyst} 3S + 2H_2O$$

A special activated alumina is used as absorbent and catalyst in the reactors. This material was selected because of its high adsorption capacity and ease of desorption of sulphur deposited on the surface.

During regeneration of a bed, the sulphur is vaporized by heating with a flow of inert gas circulated by a blower. The inert gas is cooled in a standard sulphur condenser (50 psig steam production), where liquid sulphur is condensed and removed. When desorption is completed, the bed must be cooled before going back on stream. To accomplish this, a portion of the cool inert gas is taken from the blower discharge and flowed through the bed to be cooled. Sulphur absorption and desorption results in Sulfreen being a batch or semi-continuous process. Figure 20.22 is a flow diagram of the process.

There are twenty Sulfreen units in operation. Recover levels of over 99% of sulphur with a three-stage Claus unit and a Sulfreen Unit are normal.

The tail gas from the incinerator after a unit of the above combination should contain less than 1,500 ppm of SO_2.

Sulfreen Units are generally the least costly of the tail gas processes, but do not have quite as high of a recovery as some of the processes discussed further on.

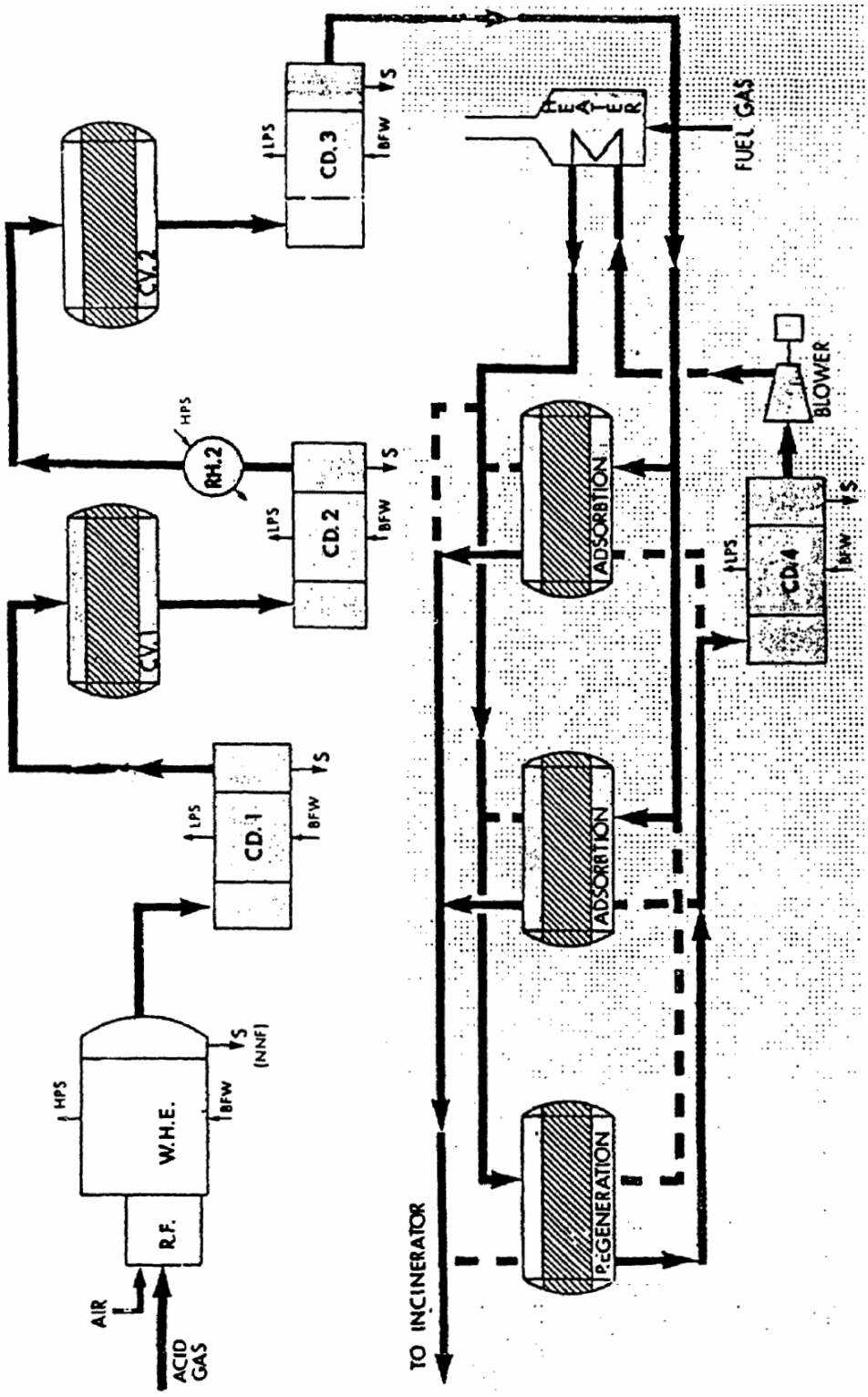


Figure 20.22: Sulfreen Flow Diagram

20.4.2.2 The Amoco CBA Process

Amoco Production Company has developed their "Cold Bed Adsorption" (CBA) process through the pilot plant stage. It is now installed in their Crossfield plant. This process is very similar to the Sulfreen process, except CBA uses a process stream indigenous to the Claus plant to accomplish regeneration of the sulphur-fouled catalyst beds in the CBA reactors (The Sulfreen process uses an external stream, circulated through a gas heater, for this purpose). As with Sulfreen, the CBA process is basically an extension of the Claus reaction over a cool bed (260ºF - 300ºF) of conventional Claus catalyst.

The same costs and tail gas quality as the Sulfreen unit can be expected with the CBA processes. Figure 20.23 shows the process flow.

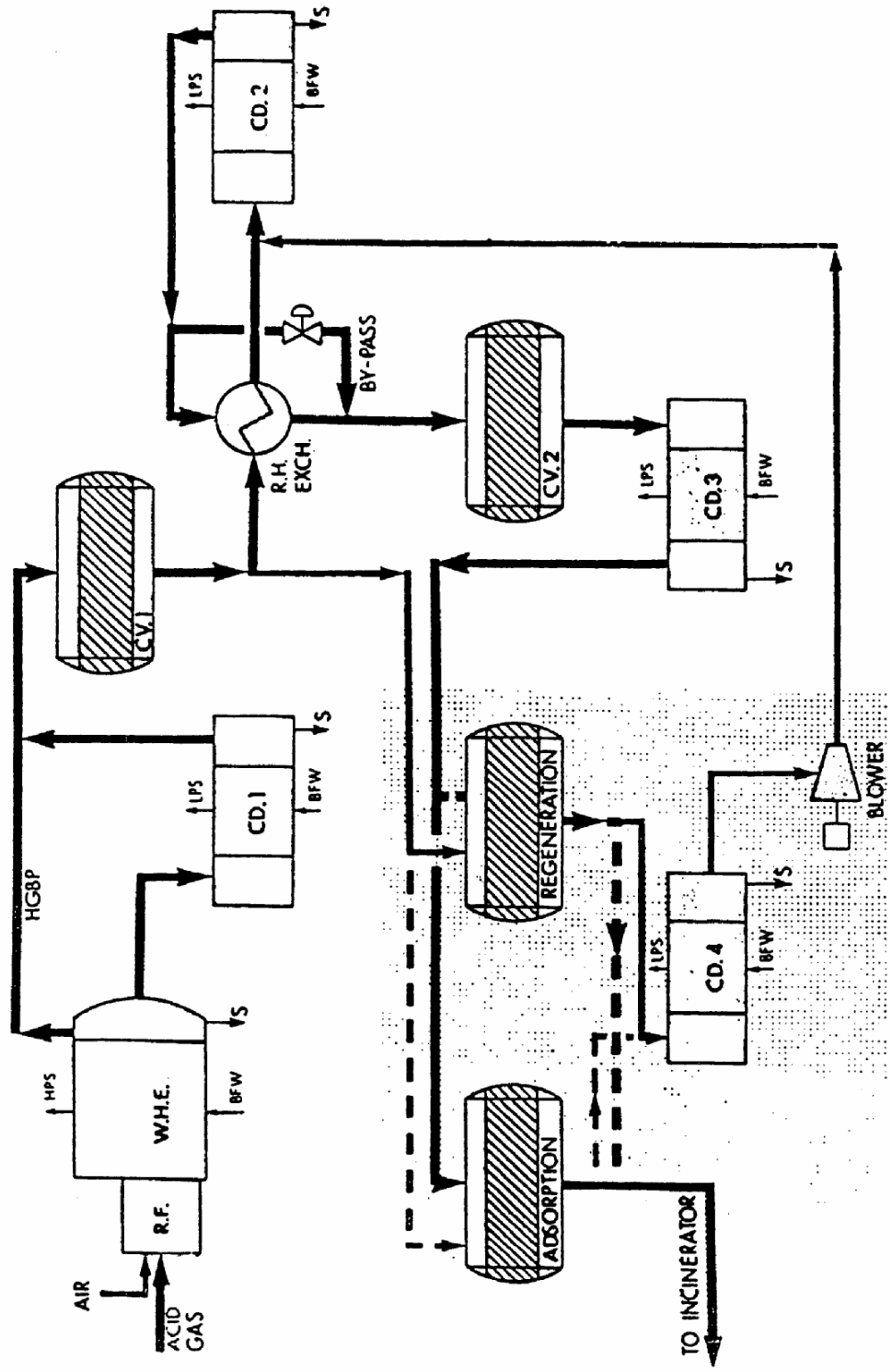


Figure 20.23: Amoco CBA Flow Diagram

20.4.2.3 The IFP Process

The Institut Francais de Petrole (IFP), of Rueil-Malmaison France, has developed their process for treatment of tail gases from Claus sulphur plants and flue gases from boilers. The process was thoroughly piloted in France; and, IFP now has twenty-five commercial plants (Claus tail gas) in operation or under construction, worldwide. Sixteen units are now in operation.

The IFP process is quite different from the processes described thus far. In the IFP process, the Claus unit tail gas, at about 260ºF, enters a vertical, packed-tower reactor vessel. The gas is contacted counter-currently by a proprietary catalyst/solvent liquid solution. The solution is recirculated around the reactor (from bottom to top), and catalyst/solvent make-up and steam condensate is added to the solution entering the reactor.

The liquid temperature in the reactor is maintained at 260ºF - 280ºF by the steam condensate injected into the solution entering the reactor. The catalyst/solvent solution has good thermal and chemical stability. The solvent is reported to be a polyethylene glycol (molecular weight - about 400) containing a metal-salt catalyst. No liquid foaming problems have been encountered. The treated tail gas leaves the top of the reactor, at about 260ºF, after incurring a total pressure drop through the system of less than 1 psi. The gas may be incinerated and flowed to a stack.

Liquid sulphur, at about 270ºF, is withdrawn from the bottom of the reactor vessel. The sulphur purity is reported to be 99.7% pure.

The basic reaction of the IFP process is the same as in the Claus unit:

Reaction 20-12

$$2H_2S + SO_2 \xrightarrow{Liquid\ catalyst} 3S + 2H_2O$$

The H_2S and SO_2 dissolve in the solvent, and in the presence of the catalyst react to form elemental sulphur. Sulphur is only slightly soluble in the solution. The IFP process does not reduce COS and CS_2 levels in the feed gas to it. While appreciable quantities of these components are formed in the Claus furnace, IFP recommends two steps to reduce the COS and CS_2 levels in the Claus unit tail gas. First, operate the first Claus converter at a hotter temperature than usual, which enhances the conversion of COS and CS_2 back to elemental sulphur vapour. Second, replace the bauxite catalyst normally used in Claus converters with a more sophisticated catalyst, which tends to promote COS and CS_2 conversion. By using these recommendations and installing an IFP plant, IFP states that an overall sulphur recovery of higher than 99.3% can be achieved. This would.probably give an SO_2 concentration of 1,000 to 1,500 ppmv in the incinerator from such a unit. The costs of the unit are estimated to be about 80% of a Claus unit associated with it.

Steam condensate is injected into the reactor to be vaporized and remove the exothermic heat of reaction of H_2S and SO_2 reacting to form sulphur. Since there is no build-up of water in the system, corrosion is essentially nil; and the unit can be fabricated of carbon steel. Figure 20.24 is a flow sheet of this process.

An IFP unit was installed in Chevron's Nevis plant. The total plant is now shut down.

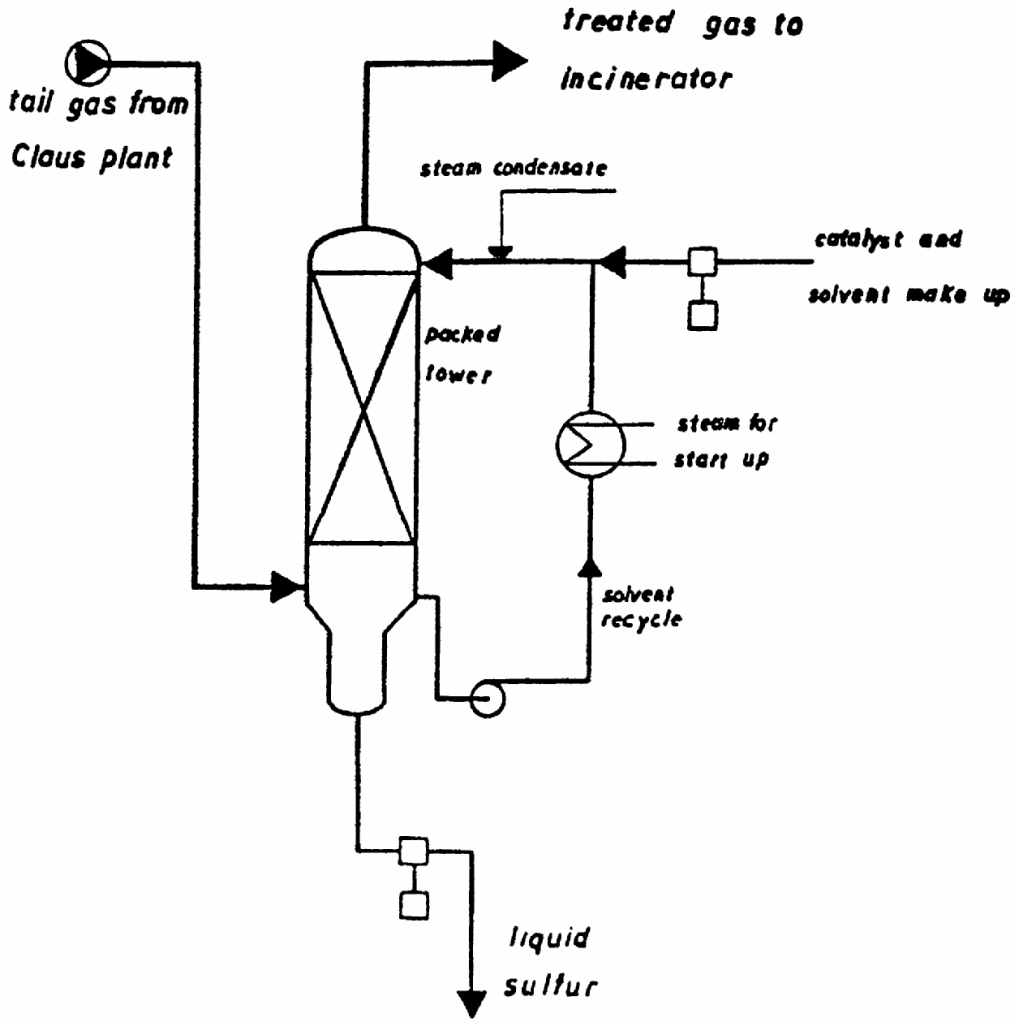


Figure 20.24: IFP Process

20.4.2.4 The SCOT Process (Shell Claus of f Gas Treating Process)

The SCOT process occurs in essentially three stages:

1. Heating and reduction.
2. Cooling and quenching.
3. H_2S absorption, stripping and recycle.

In the first stage, the Claus tail gas is heated to about 570ºF and reacted with hydrogen (or a reducing gas) over a cobalt molybdenum catalyst. All sulphurous compounds (SO_2, S, COS, CS_2, etc.) are reduced to H_2S. These reactions are highly exothermic. The hot gas from the reduction reactor is cooled in a waste heat boiler and finally quenched in a water-quench tower. The gases exit the quench step at about ambient temperature. The final stage involves the selective absorption of H_2S in an alkanolamine solution, normally DIPA. The vent gas from the SCOT absorber typically contains 200 - 500 ppmv of H_2S. This vent gas is normally incinerated before discharging to the atmosphere.

The rich amine is stripped in a conventional manner, and the H_2S-rich stream is recycled back to the front of the Claus plant. As a general rule, the DIPA solvent will co-absorb about 20-30% of the CO_2 present in the tail gas. If the feed to the Claus plant contains only 10-20% CO_2, the effect of

the inert CO_2 co-absorbed with the H_2S in the tail gas is not of any great consequence. However, if the feed to the Claus plant contains 30-50% CO_2 or greater, it is believed that the SCOT process would not normally be a good choice, due to the build-up of inerts in the Claus plant and the attendant loss in plant capacity and efficiency. Figure 20.25 is a flow sheet of the process. The cost of the unit is approximately equal to the associated Claus Unit.

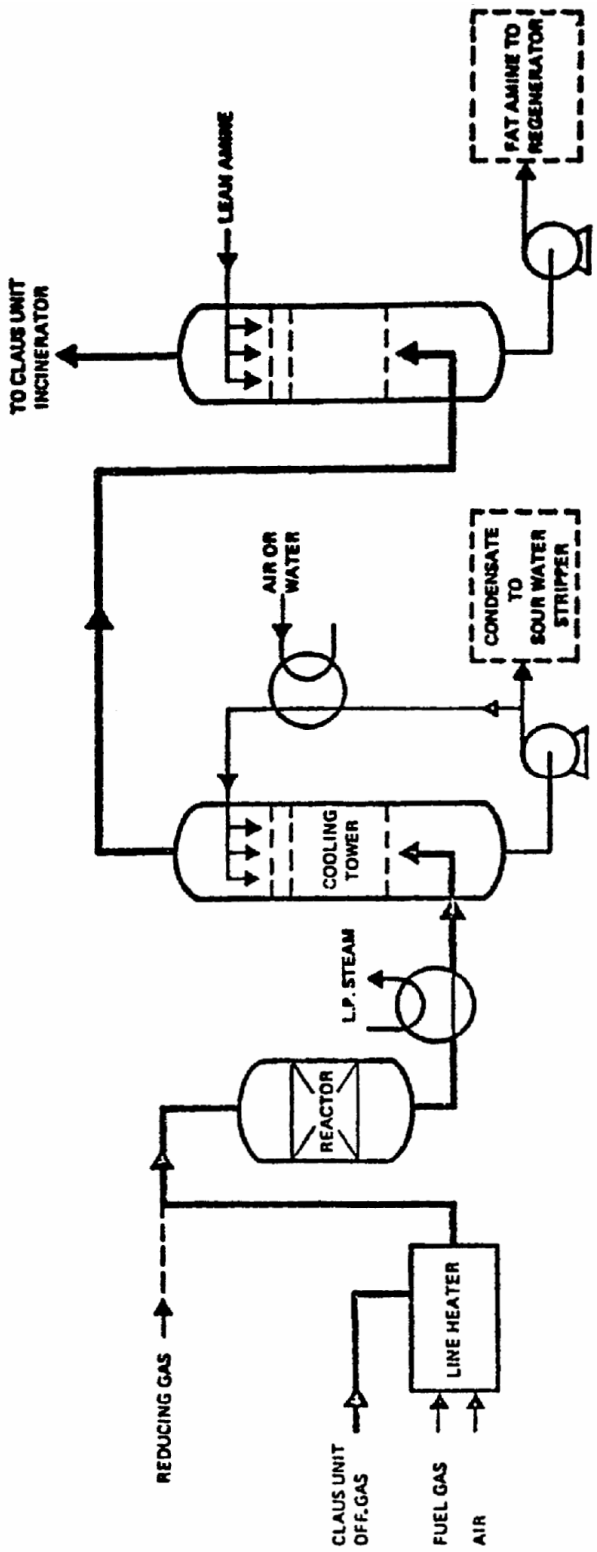


Figure 20.25: SCOT Process

20.4.2.5 The Beavon Process

In the Beavon process, the tail gas from the Claus unit is first heated by mixing with "fuel gas and air" combustion products, and then fed to the catalytic converter which contains a cobalt molybdate catalyst. The hydrogen, required f or reduction of all sulphur compounds to H_2S, is supplied by partial combustion of the fuel gas in an inline burner, which simultaneously raises the tail gas stream temperature to the level required for the hydrogenation reactions.

The converter effluent gases are then cooled with water in a direct-contact condenser, 'where most of the H_2O vapour contained in the gas is condensed, and at the same time the tail gas is cooled. The purge water produced from the condenser is of good quality, with only a small amount of dissolved H_2S. With H_2S removal in a small "sour water stripper", it is suitable for cooling tower make-up water. A recirculation loop of water with a shell-and-tube cooler is provided to maintain a flow of water into the top of the direct-contactor condenser and out of the bottom. A small bleed or purge stream of water goes to further processing.

The cooled gas then enters the Stretford absorber, where H_2S is removed almost quantitatively by the circulating Stretford solution. The treated gas from the absorber is odorless, and may be vented directly to the atmosphere or incinerated before discharge.

The Stretford process portion of the Beavon process uses an aqueous solution of sodium carbonate, which reacts with the H_2S to form NaHS, as below:

Reaction 20-13

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

The hydrosulfide is oxidized to sulphur by sodium vanadate also in the solution.

Reaction 20-14

$$NaHS + NaHCO_3 + NaVO_3 \rightarrow S + Na_2V_2O_5 + Na_2CO_3 + H_2O$$

Subsequently, the vanadium is oxidized back to the pentavalent state by blowing with air, with anthraquinone disulfonic acid (ADA) or sodium anthraquinone disulfonate, acting as an oxidation catalyst:

Reaction 20-15

$$Na_2V_2O_5 + \frac{1}{2}O_2 \xrightarrow{ADA} 2NaVO_3$$

The spent absorbent solution flows from the absorber to the regenerator (oxidizer), where air is sparged into the tower. Tiny particles of sulphur are collected as a froth at the top of the oxidizer tower. The sulphur froth is skimmed off the solution and goes down a duct to a filter or centrifuge for removal of the filtrate from the sulphur cake. The clear (regenerated) absorbent solution and the filtrate from the filter are recycled back to the absorber.

The sulphur cake may be washed, dried, melted, and added to the product from the parent Claus unit. The solid sulphur produced from the Stretford process is usually melted in an autoclave. Figure 20.26 is a flow sheet of the total process. The Stretford process flow sheet is given in Chapter 16. Reference 18 gives the latest data on the Stretford process.

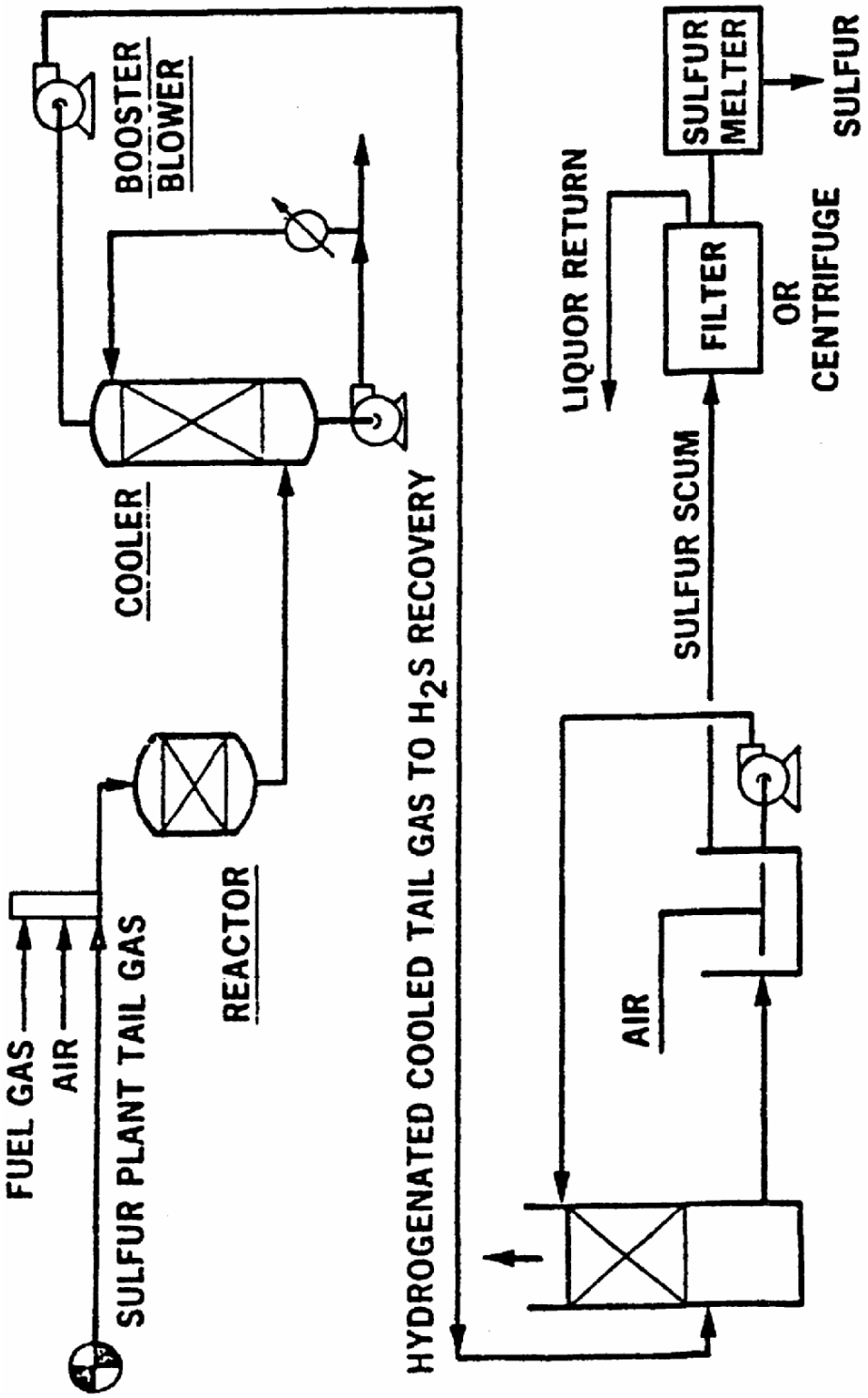


Figure 20.26: Beavon Process

20.4.2.6 The Pritchard Clean-air Process

The Pritchard Clean-air process is quite similar to the Parsons Beavon process. It also uses the Stretford process as the final treatment step. The Clean-air process also uses three processing steps. The tail gas from the Claus unit first passes through Stage III of the Clean-air process, which reduces the COS and CS_2 levels in the tail gas from about 12,000 ppmv to about 1,500 ppmv.

Stage III is actually an addition of new processing technology to the last catalytic converter in a conventional Claus unit. Therefore Stage III can easily be added to an existing Claus unit. The tail gas from Stage III flows to Stage I, which converts essentially all of the residual SO_2 in the tail gas to sulphur, with some conversion of H_2S to sulphur. Approximately 50 wt% of the sulphur recovered from the tail gas is produced from Stage I.

The tail from Stage I flows to Stage II, which is the Stretford plant. Here, essentially all of the remaining H_2S (the remaining wt% of the tail-gas sulphur production) is converted to sulphur. The tail gas leaving the Stretford plant (Stage II) contains less than 250 ppmv of H_2S, and is then generally incinerated and flowed to a stack.

20.4.2.7 The Wellman-Lord Process

In the W-L process, the gas to be cleaned is flowed to a scrubbing step where the $S0_2$ is removed by contact with an aqueous solution of sodium sulphite (Na_2SO_3) as below:

Reaction 20-16

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

The bisulphite anion (HSO_2) is found only in solution. When water is removed from a solution of sodium bisulphite, a solid called sodium pyrosulphite ($Na_2S_2O_5$) is formed:

Reaction 20-17

$$2NaHSO_3 \rightarrow Na_2S_2O_5 \downarrow + H_2O \uparrow$$

For simplicity of the chemistry involved, the composition of all solutions containing sodium bisulphite are calculated and expressed as sodium pyrosulphite. When SO_3 is present, the following reaction occurs in the scrubber:

Reaction 20-18

$$2Na_2SO_3 + SO_3 \rightarrow Na_2SO_4 + Na_2S_2O_5$$

Also, in the presence of oxygen, oxidation of the sodium sulphite solution will occur as follows:

Reaction 20-19

$$Na_2SO_3 + \frac{1}{2}O_2 \rightarrow Na_2SO_4$$

The "spent" solution from the scrubber passes to the evaporator-crystallizer in the "chemical area" of the plant, where the H_2O is evaporated from the sodium bisulphite solution. The SO_2 is also released and sodium sulphite crystals are precipitated:

Reaction 20-20

$$2NaHSO_3 \rightarrow Na_2SO_3 \downarrow + SO_2 \uparrow + H_2O \uparrow$$

Also,

Reaction 20-21

$$Na_2S_2O_5 \rightarrow Na_2SO_3 + SO_2 \uparrow$$

The composition of the solution is determined by the partial pressure of SO_2 above the liquid. Caustic (NaOH) make-up is added to the regenerated solution, which is recycled from the regeneration step to the scrubber (absorption step). The caustic reacts with the $NaHSO_3$ in the scrubber to form additional Na_2SO3 as below:

Reaction 20-22

$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O$$

Or, the reaction of caustic reacting with SO_2 in the scrubber may be written as:

Reaction 20-23

$$NaOH + SO_2 \rightarrow NaHSO_3$$

In the W-L process for Claus tail-gas treatment, the tail gas is first incinerated, then it is cooled in a waste heat boiler and by an adiabatic water-quench system to 140ºF, and finally is fed to the W-L absorber for SO_2 removal. The product SO_2 released from the absorbent in the evaporator-crystallizer is cooled and recycled to the front end of the Claus unit. The cost of this unit is about 130% of the associated Claus unit.

20.4.2.8 The Lo-Cat Process

This process is a means of removal for small amounts of H_2S from large amounts of acid gas and producing elemental sulphur. The sulphur, however, is generally not in a saleable form and is taken to a landfill.

The process consists of oxidizing the $S^=$ ion to elemental *S* by means of a metal ion usually *Fe* by the reaction indicated below:

Reaction 20-24

$$S^= + 2Fe^{+++} \rightarrow S^\circ_{(solid)} + 2Fe^{++}$$

then reactivate the Fe^{++} to Fe^{+++} by the following reaction.

Reaction 20-25

$$\frac{1}{2}O_2 + H_2O + 2Fe^{++} \rightarrow 2OH^= + 2Fe^{+++}$$

Two vessels are required, one for absorption of the H_2S arid one for precipitation of the elemental sulphur and regeneration of the ion solution.

Approximately 30 commercial installations are operating; four in Alberta. Figure 20.27 shows the process and Reference 13 describes the process.

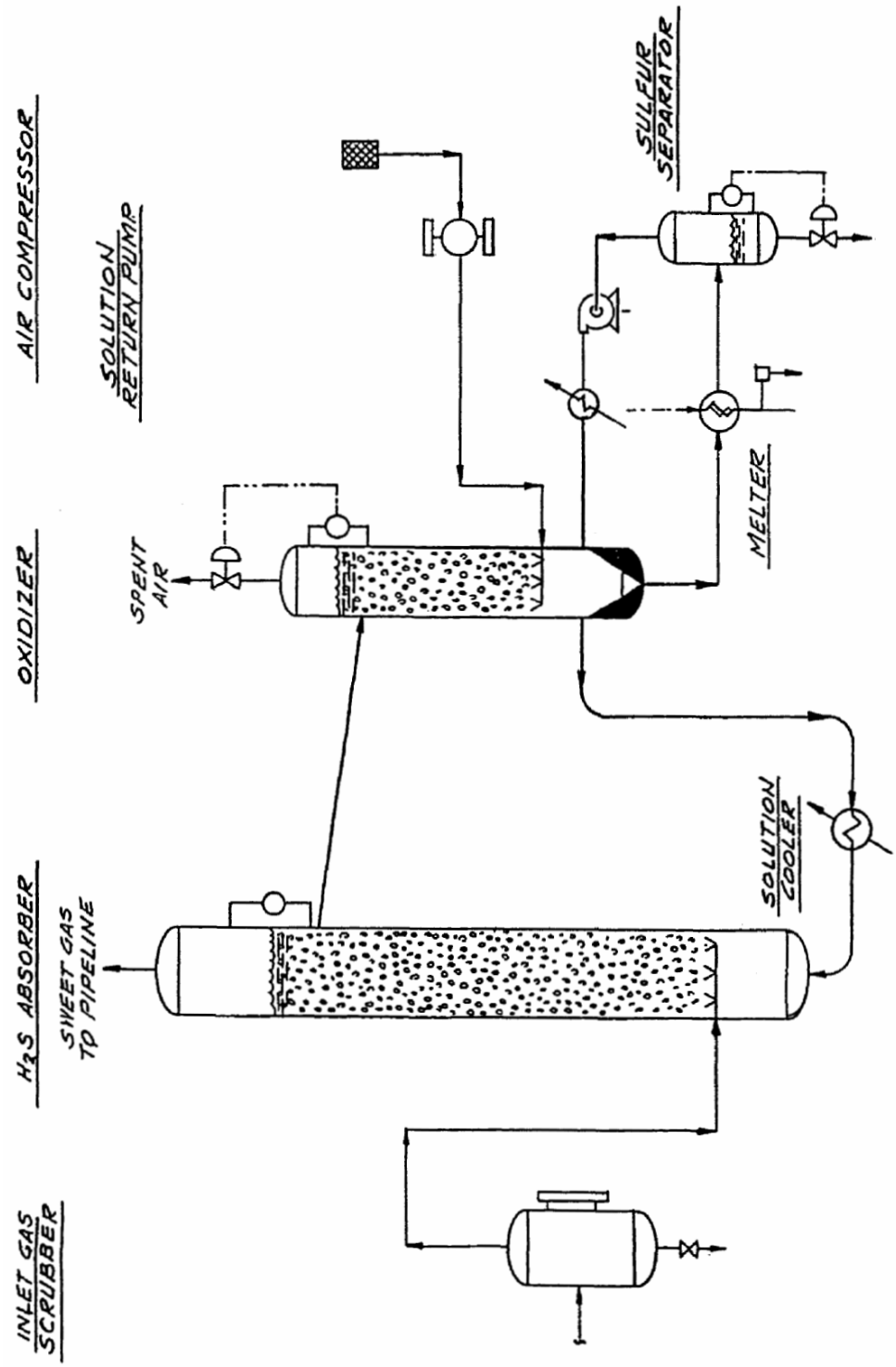


Figure 20.27: Lo-Cat Process

20.4.2.9 The MCRC Process

A unique combination of the normal Claus process and a full gas process has been installed at Pine River, B.C. for Westcoast Transmission. This process has been described in several articles. The most recent is Heigold and Berkeley.

The process consists of the regular sulphur furnace and first converter as in the Claus process and then a series of sub-dewpoint converters as in the Sulfreen process which are regenerated when the sulphur has deposited on the catalyst.

Figure 20.28 shows the process with three sub-dewpoint beds, two of which are on the line at any time.

The relative capital cost for the process and the expected recovery levels are given in the table below: converter

Type of Process	Approximate Sulphur Recovery %	Relative Capital Cost %
Two Converter Claus	92.5 - 95.8	80
Three Converter MCRC	95.0 - 97.2	105
Three Converter Claus	97.0 - 98.0	100
Four Converter Claus	98.4 - 99.0	110
Four Converter MCRC	99.0 - 99.5	125
Three Converter Claus and Sulfreen	99.0 - 99.5	140
Three Converter Claus and SCOT	99.3 - 99.7	180

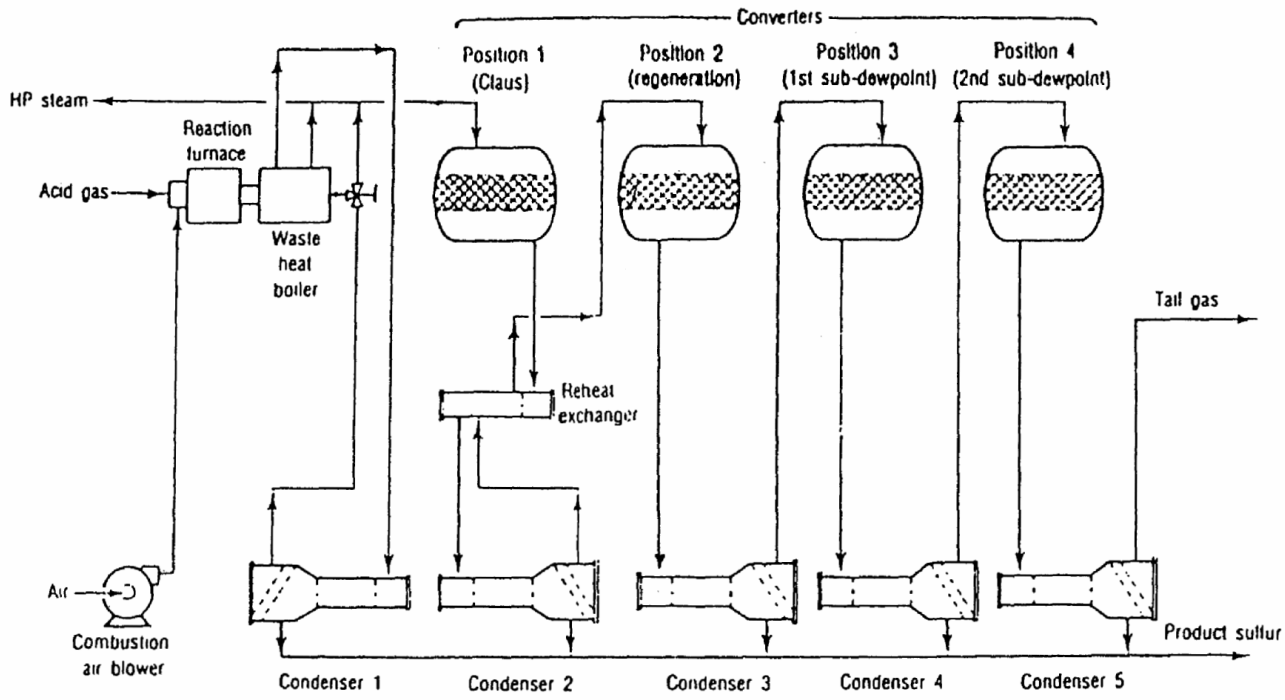


Figure 20.28: Four Converter MCRC Process

20.4.2.10 Other Processes

There are many other processes in the pilot plant or laboratory stage. Please refer to Reference 16 for further details.

20.5 Problems

Example Problem 20.1

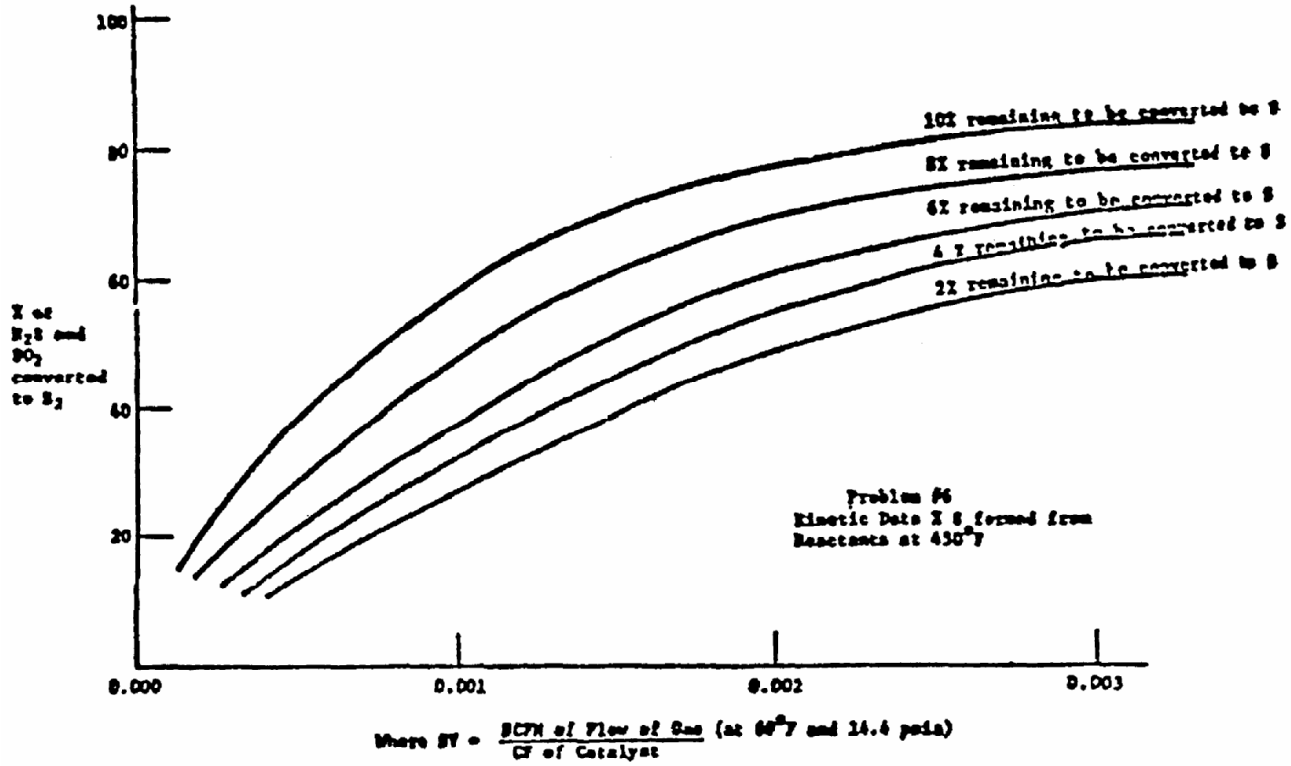
You are to determine the number of additional sulphur converters required to increase the recovery of a two-stage plant from 90% sulphur to 98%. The extra converters are to be the same size as the present ones which hold 2,000 cu. ft. of catalyst each. After each converter the necessary sulphur condenser is to be installed to remove all sulphur formed. Also a reheater will be put in prior to the next converter. Assume all reaction takes place at 450F and assume the reaction is isothermic. The kinetic data for the process is given in the attached Figure.

The feed to the sulphur plant in moles per hour is as follows:

CO_2	550
H_2S	800
H_2O	100

Assume that air is brought in the ratio such that the exact amount of SO_2 is formed for complete conversion of the H_2S to S.

Assume that sulphur is formed only as S_2 and is condensed after each converter.



Example Problem 20.2

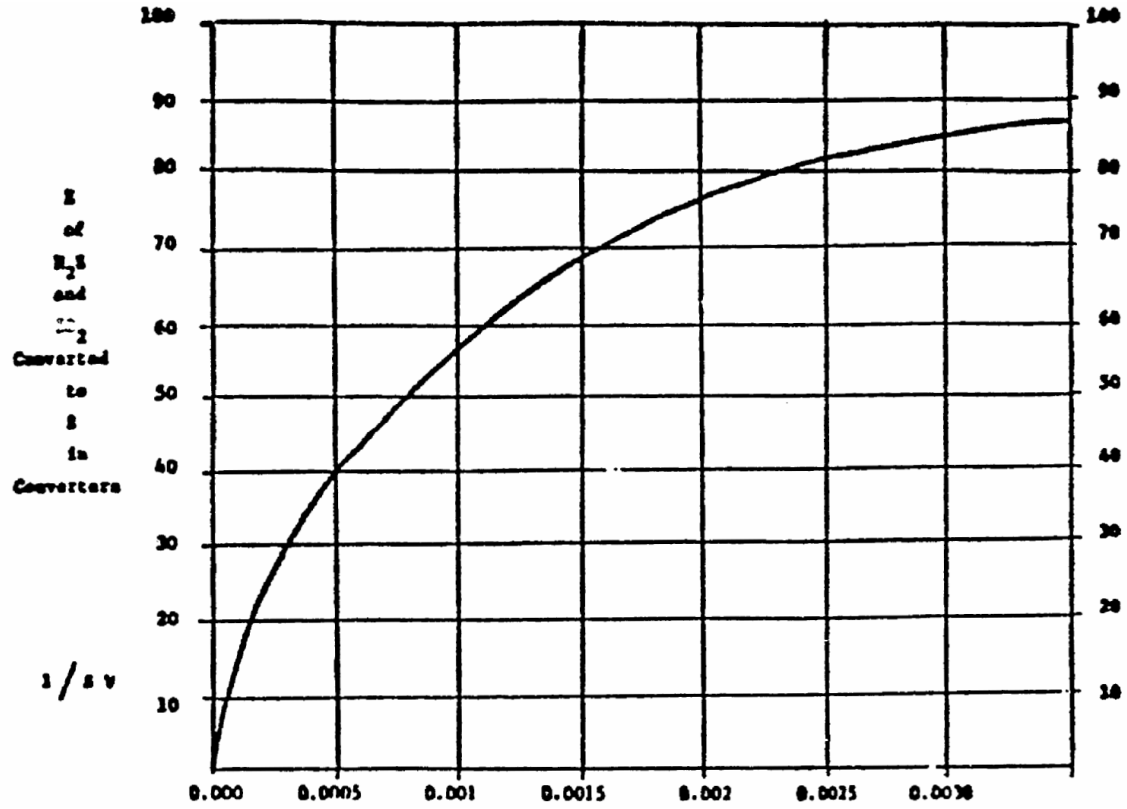
A sulphur plant that is normally fed with 60 MMSCFD of acid gas contains 80.0% of H_2S and 18.0% CO_2 and the rest is water vapour.

A bad upset occurs such that the first converter bed is coked up such that the catalyst activity is only 40% of the new catalyst and the second converter has a catalyst that has an activity of 75% of the new catalyst. The third converter has an activity of a new catalyst.

Because of a critical gas shortage the Transmission Company asks that the plant be kept on if at all possible. The governmental authorities have requested an estimate of the sulphur in long tons per day that will be discharged to the atmosphere. Please calculate this if the plant continues to run at a normal rate.

Data

- The converters each contain 8,100 cubic feet of catalyst.
- Assume 30% of the sulphur is formed in the reaction furnace.
- The attached graph gives the Approximate Conversion expected in each converter with the appropriate space velocity.
- Also for simplification the O_2 required is to completely react with the H_2S.



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21.0 Special Situations

- Solution gas recovery, cycling plants, conversation requirements, mercaptan removal.

There are many so-called special situations in gas handling or processing that have not been considered as yet. Some of the important ones are:

21.1 Solution Gas Recovery

This is recovery of treater separator vapours and stock tank vapour by compressing them, then extracting heavier components. The requirement for these types of plants is generally a fairly large oil field although recently small packaged plants can be put in individual batteries and be economical. It is important to know the state of the oil, ie. is it under flooding or being produced at formation pressure?. Also, what are the plans for the future of the field? Some of the typical plants are:

- The complex in Pembina operated by Amoco.
- IOL's Boundary Lake Plant.
- Shell's Plant at Innisfail.
- Dome's Steelman Plant in Saskatchewan.
- Dome's Sturgeon Lake Plant.
- Shell's Simonette Plant.
- Dome's Zana Lake Plant.
- Dome's Wiladen Green Plant.

The particular items to be aware of with regard to these plant are that the gas is available at much lower pressures and will have to be compressed for delivery and that it contains a much greater portion of heavy hydrocarbons (see Chapter 1).

21.2 Cycling

This is one of the oldest types of gas processing carried out. It is used primarily for wet gas fields with a high quantity of liquids. The process is used primarily to recover the liquids. It consists of producing the gas and condensate together separating the condensate and repressuring the gas and injecting the gas back into the formation. By keeping the reservoir pressure up,. the recovery of liquids is greatly improved. A typical liquid recovery curve is as follows:

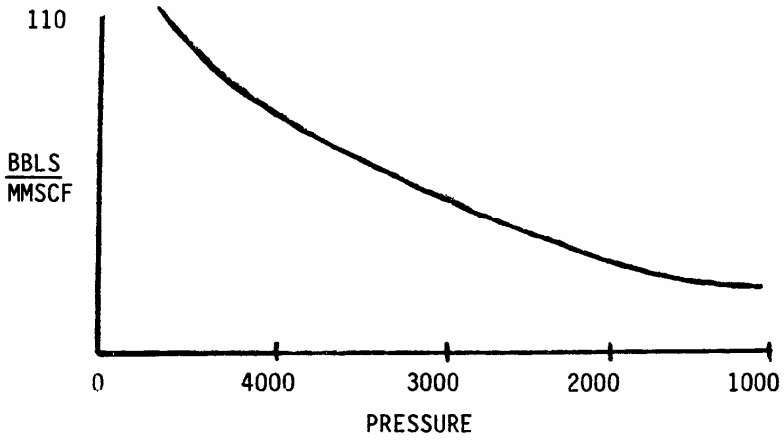

Figure 21.1: Liquid Recovery with Blowdown

The problem of dropping the pressure is the phenomena of retrograde condensation which was considered early in the book and is shown by Figure 21.2 (see Chapter 3). The dropping of pressure in the formation causes liquids to drop out in it and reduces the recovery.

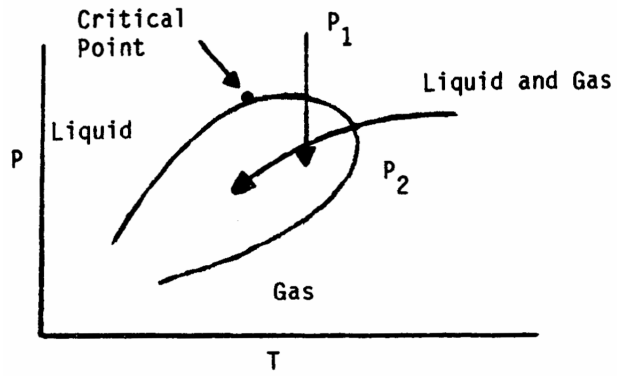

Figure 21.2: Pressure Drop - Retrograde Condensation

The pressure is kept up for one complete cycle so that the reservoir is not in the two phase area during this cycle.

Normally, cycling is not considered economical when C_5^+ content is less than 20 to 25 BBLs/MMSCF. Some examples of cycling plants are:

- Harniatton Elkton operated by Mobil.
- Carson Creek operated by Mobil.
- Kaybob (all three plants) operated by Dome and Chevron.
- Whitecourt (Windfall) operated by Amoco (this is not exactly cycling as another field is being used for repressuring).
- West Pembina Plant operated by Dome.

Generally, the Conservation Board always asks if cycling of the field has been examined whenever a new project is put before it.

21.3 Load Levelling

A similar type of facility or scheme to recycling is load levelling. The gas companies have a varying load, thus they have a larger demand in the winter than summer or spring. The two major gas buyers have demands generally as follows:

	TC	A&S
Yearly contract volume	1.00	1.00
Agreed minimum take	1.00	0.90
Maximum daily demand	1.33	1.20
Normal minimum but can go lower	0.75	0.75

Thus, in a field where the by-products are quite valuable, a scheme can be developed that keeps the plant going at capacity and compresses excess gas into reservoir while still extracting by-products. Some increase in overall liquids and other by-products recovery occurs with this scheme. Figure 21.3 shows the process for load levelling.

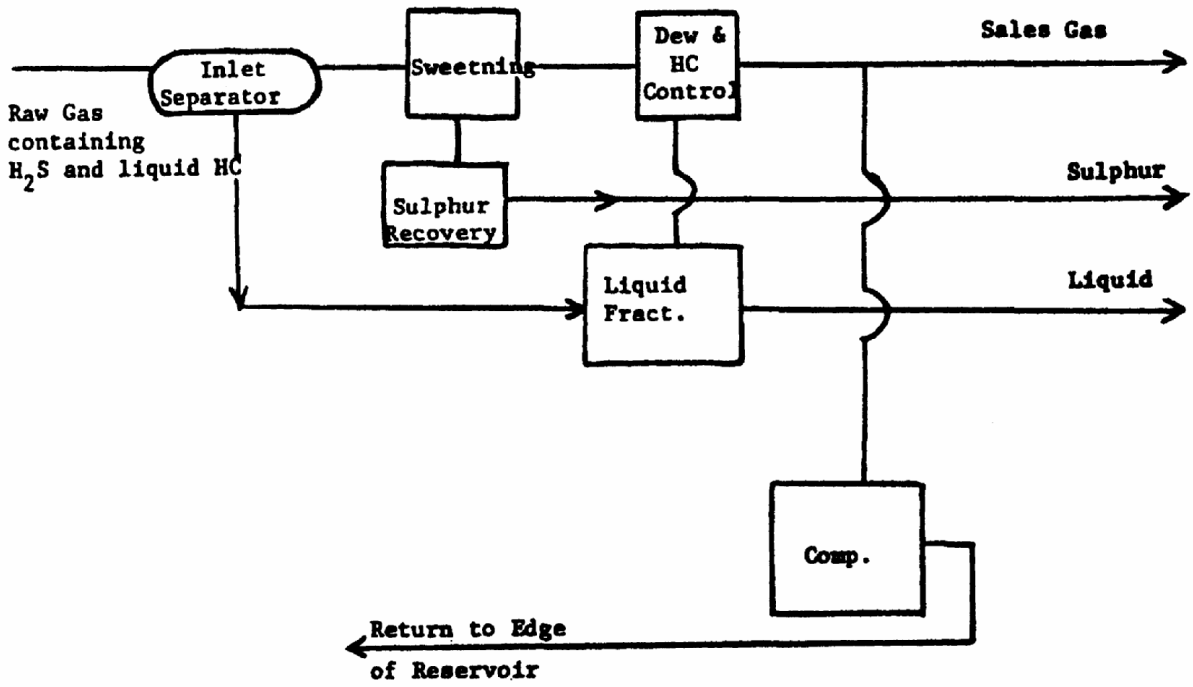

Figure 21.3: Load Levelling Process Scheme

21.4 Enhanced Oil Recovery Projects (EOR Projects)

There has recently developed a great deal of interest in greater oil recovery using miscible fluids. The primary miscible fluids used are either a mixture of ethane and propane or carbon dioxide (CO_2). The ethane/propane mix is produced as explained in the LPG recovery chapter and CO_2 is obtained from CO_2 wells (in the United States only) or from the tail gas from sulphur plants, ammonia plants, power plants, etc. (This is in the pilot plant stage only).

Another material that is used as a fluid for oil recovery is nitrogen. This is produced by cryogenic processing of air.

The recovery or removal of these miscible fluids during or after the production of the oil will generally follow the normal solution gas processing methods.

21.5 Mercaptan Removal

Another problem is mercaptan removal. This is a problem when trying to meet A&S specifications. The specifications for mercaptan are now 0.2 grains/100 CF rather than 5 to 10 grains/100 CF. Mercaptans can be removed by:

1. Deep LPG extraction; CH_3SH comes out with C_3 and iC_4. The equilibrium "K" value for CH_3SH (methylmercaptan) is approximately 1.10 of the value for iC_4.
2. Caustic washing with 10% NaOH.
3. Some patented sweetening processes.
4. A special solvent in the treating solution using a material in the regular treating solution called tretaglyamine see reference 8.
5. Molecular Sieves.

As mentioned above, the mercaptans and other impurities of a similar nature are often removed from the gas when LPG's are recovered and are in the LPG's. These impurities have to be removed before propanes, butanes, etc. can be sold as specification products. Three references (4,5,6) cover this subject in some detail. The primary processes used for the removal of mercaptans are molecular sieve treating, regenerative caustic treating, and merox treating. Figures 21.4, 21.5 and 21.6 show these processes.

In the molecular sieve process, the liquid is passed through a bed of molecular sieves selected for the compounds that have to be removed until the sieve is saturated with these compounds. It is then taken off the line and regenerated by heat. The normal cycle time for a unit is 6 - 12 hours. The sieve generally removes compounds in the following order:

1. water.
2. methanol.
3. lower molecular weight mercaptans.
4. higher molecular weight mercaptans.
5. hydrogen sulphide.
6. carbonyl sulphide.

This is a very good treating process, but it uses a considerable amount of fuel on the regenerative step. Figure 21.4 is a flow sheet of the process.

The regenerative caustic process uses a 10% NaOH stream for removing the mercaptans. Only mercaptans can be removed on a regenerative basis; the other sulphur compounds like H_2S and COS are removed on a batch basis in a pre-treatment tower.

This process is quite corrosive in the regenerative stage and difficult to operate. The product from the process has to be dried in a separate unit. Figure 21.5 is a flow sheet of the process.

A proprietary process utilizing principles of the caustic system is called the Merox Process.

Figure 21.6 shows a simplified flow sheet of this process which is installed at Dome's Sarnia Plant. The Merox Process developed by UOP is a catalytic chemical treatment to convert mercaptans to disulfides at normal operating temperatures of 80ºF - 120ºF. Sour butane is fed to the extractor

where in the same manner as in the caustic system, mercaptans are removed by the caustic from the butane. The rich caustic which has the Merox catalyst in solution is mixed with air and the mercaptans are converted to disulfides in the oxidizer. The disulfides are separated off in the dilsufide separator and the lean caustic solution containing Merox catalyst is recirculated back to the extractor. The disulfides are generally just sent to the hydrocarbon condensate system.

This system uses much less fuel and is not nearly as corrosive as the regular regenerative caustic system.

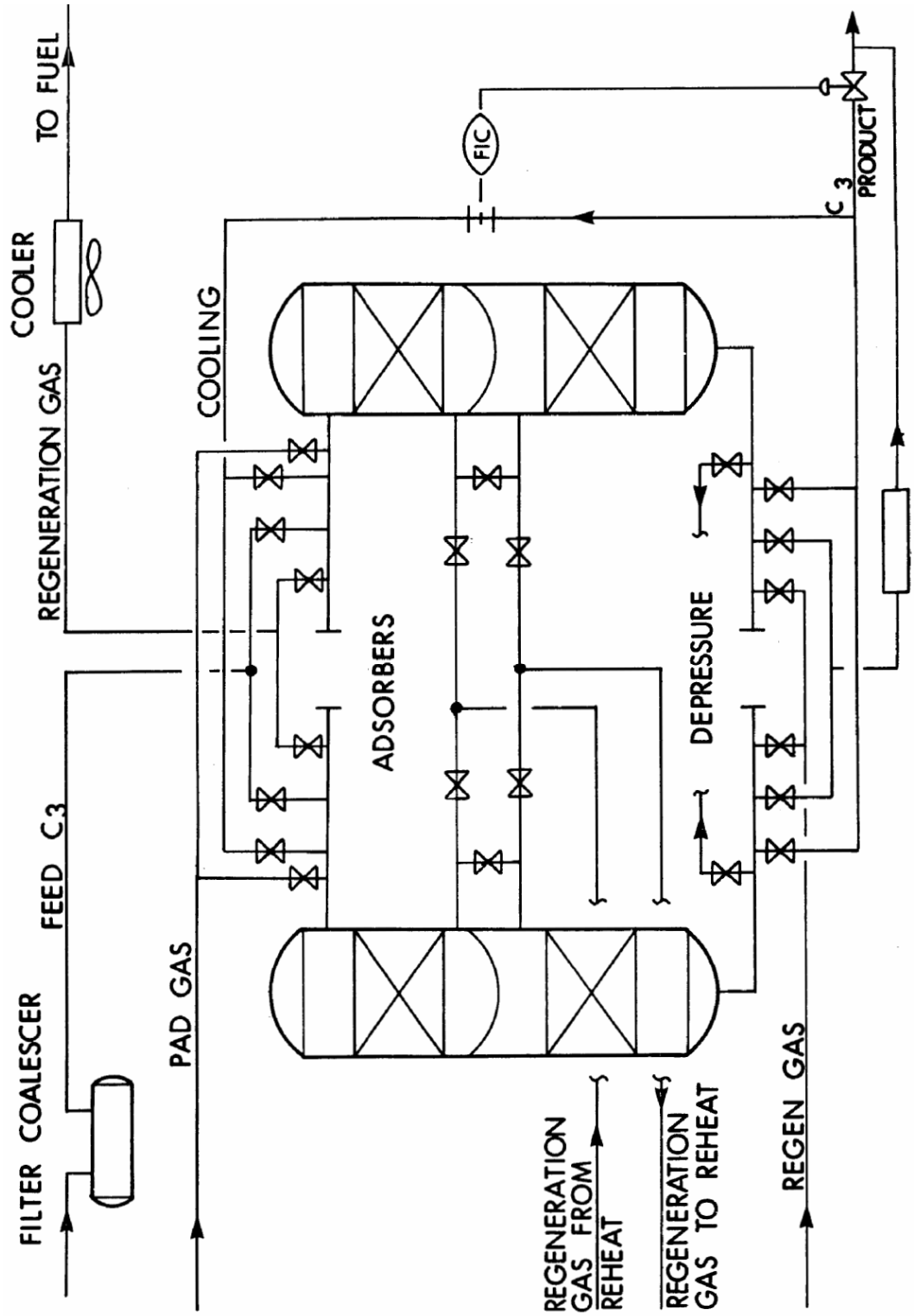

Figure 21.4: Moleculer Sieve Treating & Drying

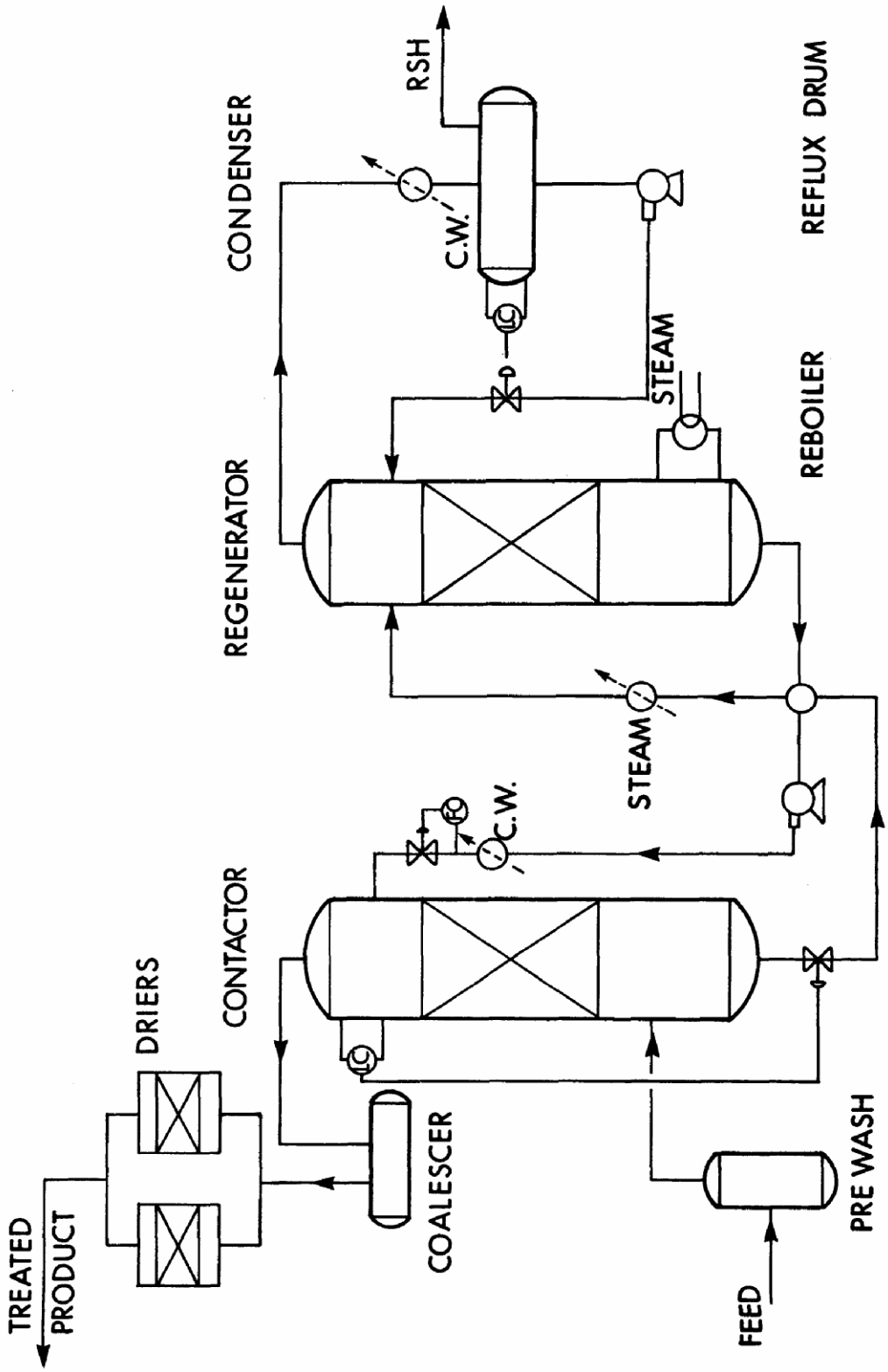

Figure 21.5: Regenerative Caustic

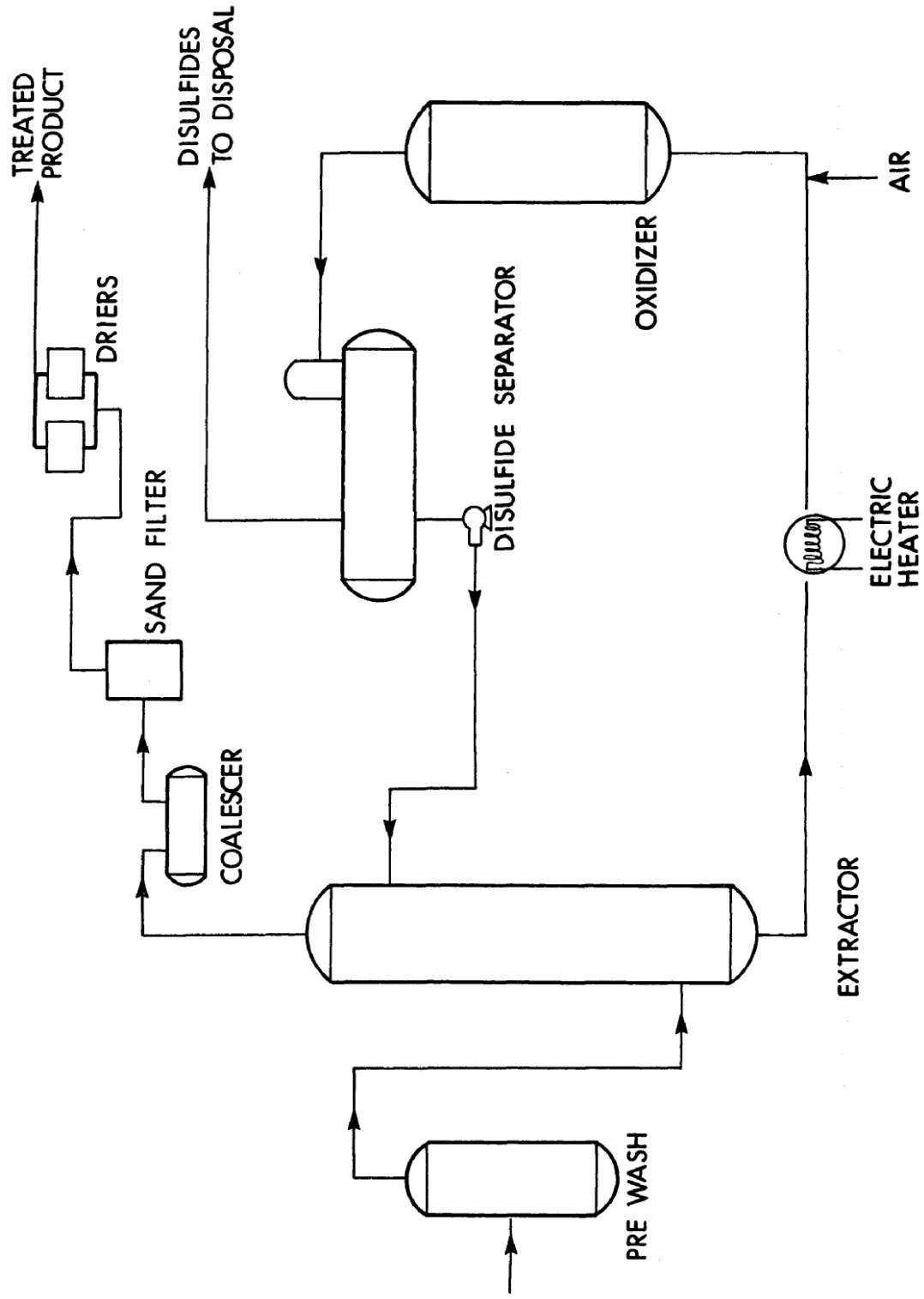

Figure 21.6: Merox Extractor

21.6 Gas Processes

There are many and varied gas processes used in the processing of natural and substitute gas. Hydrocarbon Processing every two years comes out with a list of current processes along with a description, simplified flow sheet, and a list of typical installations. Table 21.1 is the list for 1984.

Table 21.1: Processes for Natural and Substitute Gas

Drying

High concentration glycol dehydration 78
Drizo dehydration 80

Treating

Benfield 79
Activated MDEA 80
ADIP 80
Amine guard 80
Amisol 80
Beavon sulfur removal—MDEA process (BSR/MDEA) 80
Catacarb CO_2 and H_2S removal 81
Catasulf 81
Chemsweet 81
Clauspol 1,500 81
DEsulf H_2S removal 81
DIAMOX 81
FLEXSORB absorbents 82
Fluor improved Econamine 82
Fluor solvent 82
Giammarco-Vetrocoke 82
Mercury removal (Calgon) 82
Mercury removal (IFP) 82
Optisol 82
Potash vacuum process 83
Purisol 83
Rectisol 83
Selefining 83
SELEXOL 83
SNEA (P)—MDEA 83
SNPA-DEA low energy 85
Stretford process 85
SULFINOL 85
Sulfint 85
Sulfur—H_2S removal by activated carbon 85

Recovery

CO_2 recovery 84
EOR nitrogen rejection 85
Gas/Spec FT technology 85
Nitrogen rejection/HC recovery 87
LPG recovery from refinery offgases 87
PetroFlux 87
Ryan/Holmes technology 87

Sulfur

Amoco sulfur recovery 86
Autocirculation Lo-Cat H_2S oxidation 87
Catalytic sulfur degassing 88
Lo-Cat hydrogen sulfide oxidation 88
Parsons Claus sulfur recovery 88
Selectox and Recycle Selectox 88
SNEA (P) catalytic sulfur degassing 88
SNEA (P) sulfur recovery 88
SRU 89
Sulfur degasification 89
Sulfur degassing 89
Sulfuric acid from alkylation sludge 89
Topsøe Wet Gas Sulfuric Acid Process 89
Unisulf 89

NOTE: **Bold face** denotes process with flowsheet.

Effluent cleanup

Shell Claus offgas treating process (SCOT) 90
Beavon Sulfur Removal Process (BSRP) 91
BSR/Selectox 91
Catalytic incineration 91
CBA 91
COPE 91
Econ-Abator catalytic incineration 91
FLEXSORB SE absorbent 93
Selectox-catalyzed incineration 93
SULFREEN 93
Sulften 93
ULTRA 93

Hydrogen

Hydrogen purification—Polybed 92
Cryofining 93
Cryogenic purification 94
Hydrogen 94
Hydrogen by methanol decomposition 94
Hydrogen-Prism 94
Hydrogen-Topsøe 94
REMET 94
Steam reforming 97

Liquefied natural gas

MCR liquefaction 95
Expander liquefaction 97
Tealarc 97

Flue gas cleanup

SOLINOX 96
NKK NO_x removal 97
NKK SO_2 removal 97
Stackpol 150—SO_2 97
Thermal DeNO$_x$ 99

Liquid gasification

Lurgi—pressure gasification of coal 98
Autothermal reforming 99

Gasyntnan 99
GI two-stage gas producer 99
High-temperature Winkler gasification 99
HYTORT process 99
JGC methane-rich gas (MRG) 99
Koppers-Totzek coal gasification 100
PRENFLO coal gasification 100
RENUGAS 100
Shell coal gasification—SCGP 100
Shell Gasification 100
SPARG-Process 100
Texaco coal gasification 102
Texaco partial oxidation 102
U-Gas 102

CH4 Shift & methanation
Sulfur tolerant shift (SSK) 101
IRMA methanation 102
Methanation 102

The HMProcess 105
SNG from CO-rich gases 105
Syntar process 105

Liquid treating
Merox 103
Bender 105
Merifining/Napfining/Mericat/Thiolex/Regen 105

C3-C5 C3—C5 conversion
Butamer 104
Butene-1 recovery (Sorubtene) 105
CATADIENE 106
Catofin 106
C4 hydroisomerization 106
C4 isomerization 106
C5 isomerization 106

21.7 Substitute Natural Gas

21.7.1 What is SNG?

SNG means substitute natural gas, synthetic natural gas, or supplemental natural gas, which is produced from heavier, liquid hydrocarbons or coal. The resulting gas is essentially methane (over 98%) with some amounts of carbon dioxide, hydrogen and carbon monoxide present, and has a gross heating value of about 1,000 BTU's/CF. SNG is a complete substitute for common natural gas and is mixed with it in transmission lines, distribution lines and combustion equipment. During the seventies, in North America, when there was a perceived gas shortage, considerable interest was generated in this method of supplying gas and several plants using liquid feed stocks were built and several demonstration plants using coal were also built. Many of these plants have been shut down. However, in places where natural gas is not available (ie. South Africa), these types of plants provide a good source of methane, and because of North America's extensive coal and heavy oil reserves (tar sands, shale oil, etc.), they can provide an alternate to very expensive gas from arctic areas or very deep, tight wells.

21.7.2 How is SNG Made?

The basic chemical reactions in making SNG are as follows:

a) For oil based feed stock

Reaction 21-1

$$4(CH_2) + 2H_2O \rightarrow 3CH_4 + CO_2$$

b) For coal

Reaction 21-2

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

c) For methanol

Reaction 21-3

$$4CH_3OH + 3CH_4 \rightarrow CO_2 + 2H_2O$$

As can be seen, the basic process is to convert oils or coal into methane. The less hydrogen is present in the mix, the more difficult is the processing. A measure of the difficulty is the carbon to hydrogen weight ratio in the feed stock. The approximate capital costs (in 1971) for making SNG for various feed stocks is given in Figure 21.7. The abscissa shows the C/H weight ratio for the feed stocks. As can be seen, the cost for a coal feed stock plant is about four times as great as the cost of an LPG feed stock plant. In 1975, the cost was estimated to be about 6 times as great as an LPG feed plant.

Another consideration in the type of feed is whether the feed can be fed directly after treating to a catalytic unit to react with steam and produce SNG. The materials with low boiling ranges can; others, such as crude oil, residual oils, etc. have to be hydrogenated prior to gasifying. A similar and still more complex process is required to hydrogenate and gasify coal.

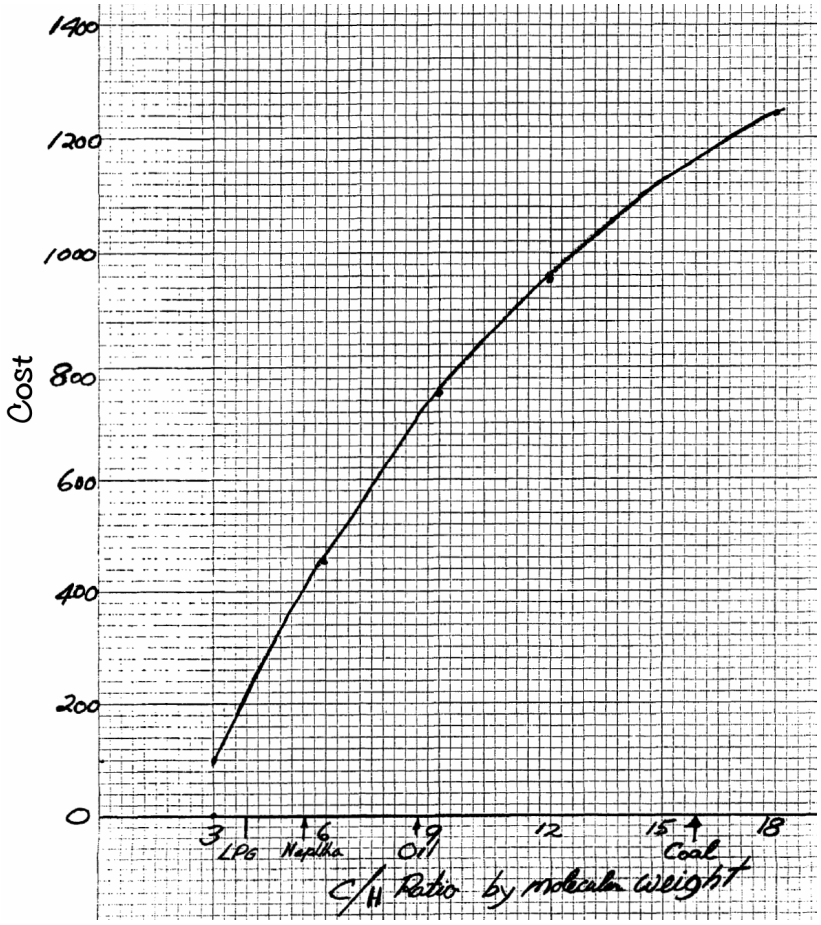

Figure 21.7: Cost of SNG Manufacturing 1971

21.7.3 Plants in Operation

The process of liquid hydrocarbon and coal gasification has been carried out away from North America for many years.

Plants using naphtha as a feed stock have been built and are in operation in Great Britain, Japan, and Italy and more recently in the United States.

A plant for gasifying coal has been operating in South Africa for many years. Much smaller plants, essentially pilot plants, have been in operation in Germany and Scotland for some time, and several 1 MMSCFD demonstration plants are in operation in the United States.

In the United States, there were many plants using liquid hydrocarbons in operation. For example, the Columbia Gas Plant at Green Springs, Ohio produced at a rate of over 200 MMSCFD. Table 21.2 shows the data on plants constructed in the United States. Most of these have now been shut down.

Table 21.2: U.S. SNG-From-Oil Facilities

Company	Feedstock Volume, 1000 bbl/day	Feedstock Type	Feedstock Source	SNG MMcf/d	Days Operate Per Year	Est. Gas Price $/MMBtu	Cost $ — Million	SNG Process	Contractor	Plant Location
OPERATIONAL PLANTS AND PLANTS IN START-UP OR TESTING										
Algonquin SNG Inc.	25	Naphtha	25% domestic; 75% imported	120	300	$2.50—$3.25	60	CRG	The Lummus Co.	Freetown, Mass.
Ashland Oil, Inc.	11	Naphtha	Ashland's Tonawanda refinery	60	300	$2.50—$2.90	13 (plant only)	CRG	Foster Wheeler Corp.	Tonawanda, N.Y.
Boston Gas Co.	10	Propane	Imported, Exxon	60	180	$1.50—$2.25	7	MRG	The Badger Co. Inc.	Everett, Mass.
The Brooklyn Union Gas Co.	13.7	Naphtha	Exxon	60	180	$2.25—$3.00	45	CRG	The Lummus Co.	Brooklyn, N.Y.
Columbia LNG Corp.	70	Natural gas liquids, propane	30,000 bbl/day propane domestic; 40,000 bbl/day imported, Canada	250	365	$2.40—$2.80	45	CRG	Davy-Powergas	Green Springs, Ohio
Commonwealth Natural Gas Corp.	7.35	Propane, butane	Imported from Libya	30	30—180	$2.50+	67	CRG	J. F. Pritchard and Co.	Chesapeake, Va.
Consumers Power Co.	60	Natural gas liquids	Imported, Canada	280 (2 trains)	365	$2.00—$3.00+	180	CRG	The Lummus Co.	Marysville, Mich.
Gasco, Honolulu, Hawaii	3.4	Naphtha	Imported via affiliated refinery	15	365	$2.00+	7.2	Lurgi; Gasynthan	Ralph M. Parsons Co.	Barbers Point, Oahu
Hi-Gas Supply, Inc. (Northern Illinois Gas Co.)	35	Natural gas liquids	Mapco, others delivered by Mapco Pipeline	165	365	$2.00+	80	CRG	Bechtel Corp.	Morris, Ill.
Public Service Electric and Gas Co.	4	Naphtha	Domestic	20	300	$2.00—$3.00	12	CRG	Foster Wheeler Corp.	Harrison, N.J.
Public Service Electric and Gas Co.	25	Naphtha	Imported	125	300	$2.00—$3.00	30	CRG	Foster Wheeler Corp.	Linden, N. J.
Peoples Gas Light and Coke Co.	33	Naphtha	13,000 bbl/day Union; 20,000 bbl/day Amoco	160	365	?	77	CRG	M. W. Kellogg	Elwood, Ill.
Baltimore Gas and Electric Co.	12	Naphtha	Virgin Islands	60	180	?	25	Lurgi; Gasynthan	Stone & Webster	Sollers Point, Md.

21.7.4 Process Description

21.7.4.1 Liquid Hydrocarbons up to and including naphthas

This is the process that is being used primarily for new plants being built in the United States.

A process flow diagram is shown in Figure 21.8. The process is generally as follows:

The feed stock is hydrodesulphurized to a low sulphur content (less than 5 ppm), steam is added, and the mixture then heated to about 840ºF where the Reaction 21-1 takes place over a nickel based catalyst. The reaction is exothermic and the products must be cooled prior to proceeding on to the next step, which is just the further reaction of hydrogen with the remaining CO and CO_2. Table 21.3 gives the gas composition after each stage in the process.

The final stage in the process is the removal of CO_2, which seems to be primarily done by hot potassium carbonate absorption.

A slightly different version of the process that is gaining favour is the Hydrogasification process in which the feed of naphtha is split and gasified with rich gas from the first gasifier. This process is particularly applicable to light naphthas. It operates at a lower overall steam ratio and the catalyst life is reported to last longer. Figure 21.9 shows this process.

Generally, the process is run at the pressure that is required to deliver gas to the pipeline. The lower the temperature the CRG reaction is carried out at, the better for an equilibrium point of view as it prevents the formation of hydrogen.

The lighter the feed stock, the longer the catalysts last. Also, straight chain compounds are better than ring compounds.

The catalyst is poisoned by both sulphur compounds and chloride compounds, so these should be kept to a minimum. The pre-treatment section of the plant is often the most expensive section of it.

Table 21.3: Typical Product Gas Compositions from the CRG Double Methanation SNG Process

Operating Conditions:				
Process stage	**Out CRG**	**Out 1st Methanation**	**Out 2nd Methanation**	**SNG**
Gas composition mole %				
CH_4	61.2	74.8	79.0	98.05
H_2	17.0	4.6	1.1	1.45
CO	1.0	0.2	< 0.1	< 0.1
CO_2	20.8	20.4	19.9	0.5
Heating Value Btu/ft^3 (dry)	678	773	803	998
Specific Gravity (air = 1.0)	0.677	0.727	0.747	0.552
Wobbe No.	824	907	929	1343
Thermal Efficiency %	--	--	--	93
Working Pressure psi	--	--	--	375

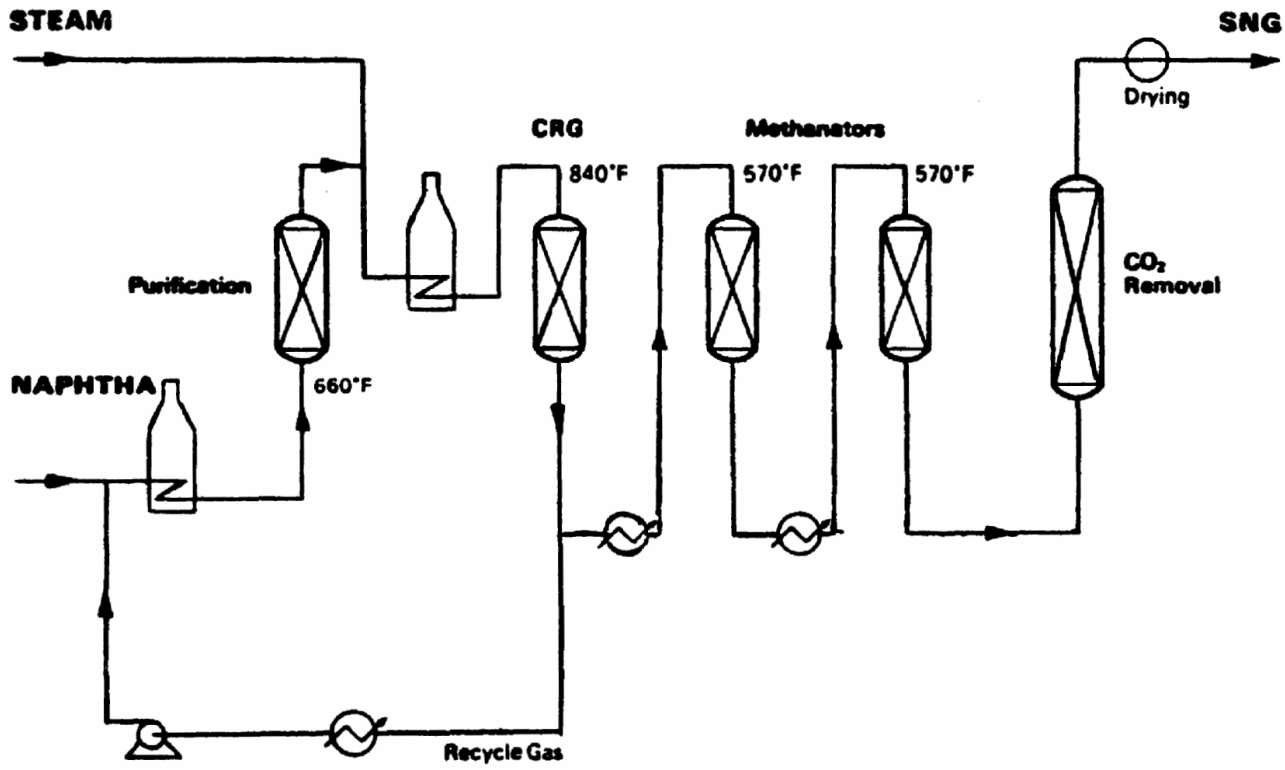

Figure 21.8: The CRG Double Methanation SNG Process

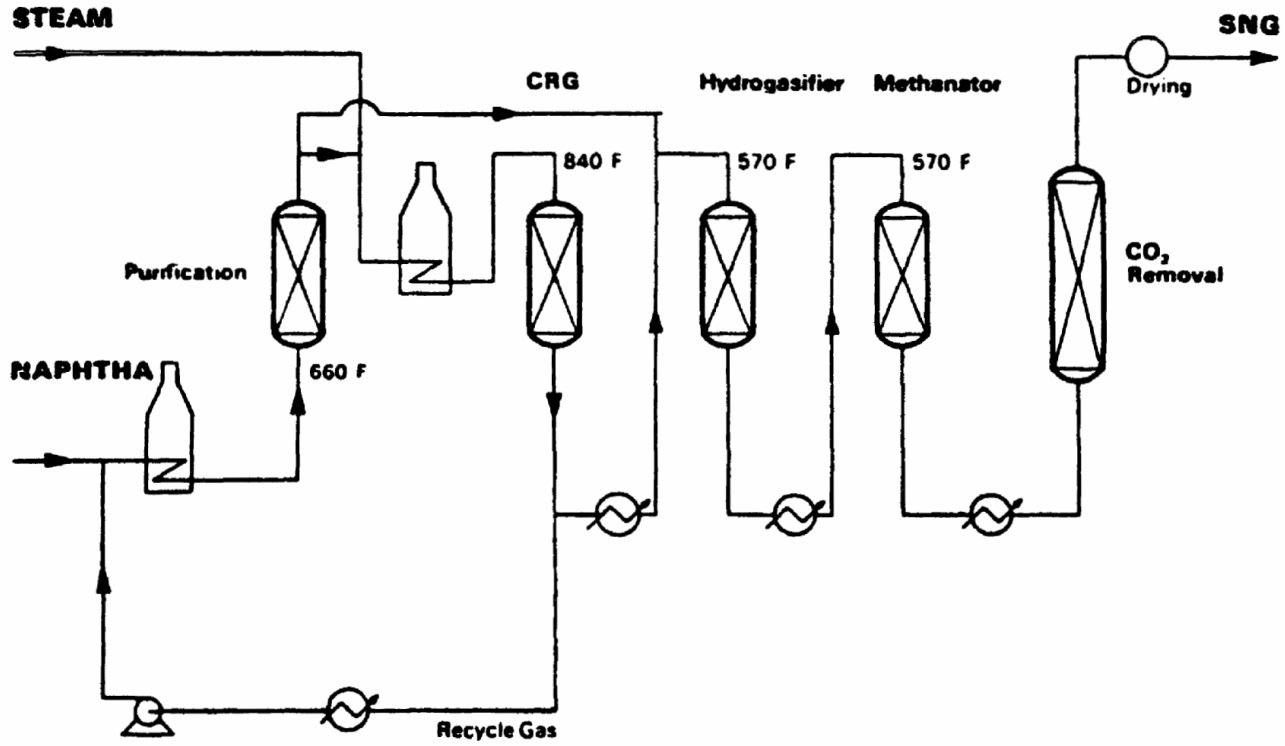

Figure 21.9: The CRG Hydrogasification Process

21.7.4.2 Heavier Liquid Hydrocarbon Feed Stocks

The process for heavier feed stocks such as crude oil requires the hydrogenation of the feed stock. The reaction that occurs is as follows:

Reaction 21-4

$$(CH_2) + H_2 \rightarrow CH_4$$

The hydrogen is produced by catalytic reaction of the product as follows (steam reforming method):

Reaction 21-5

$$CH_4 + H_2O \rightarrow 4H_2 + CO_2$$

or by partial oxidation method:

Reaction 21-6

$$(CH_4) + O_2 \rightarrow 2H_2 + CO_2$$

This process is shown in Figure 21.10. The feed stock is heated to over 1000ºF and the pressure is set at 800 psig. The feed stock passes in the fluid bed hydrogenator, the product gas is sent to a purifier and then further hydrogenated. Table 21.4 shows some types of gas compositions.

Table 21.4: Typical Product Gas Compositions from the F.B.H. Based SNG Process

Feedstock	Algerian Crude	Heavy Naphtha
Reactor pressure (lbs/in^2)	**750**	**720**
1	**2**	**3**
SNG compositions mole %		
CH_4	79.5	71.3
C_2H_6	8.7	16.4
H_2	10.0	9.1
CO	1.5	1.8
CO_2	0.2	1.3
N_2	0.1	0.1
Heating Value, BTU/ft^3 (dry)	997	1046
Specific Gravity (air = 1.0)	0.556	0.608
Wobbe No.	1336	1340
Aromatic Condensate 1b/100 lb, feedstock supplied	22.6	16.7

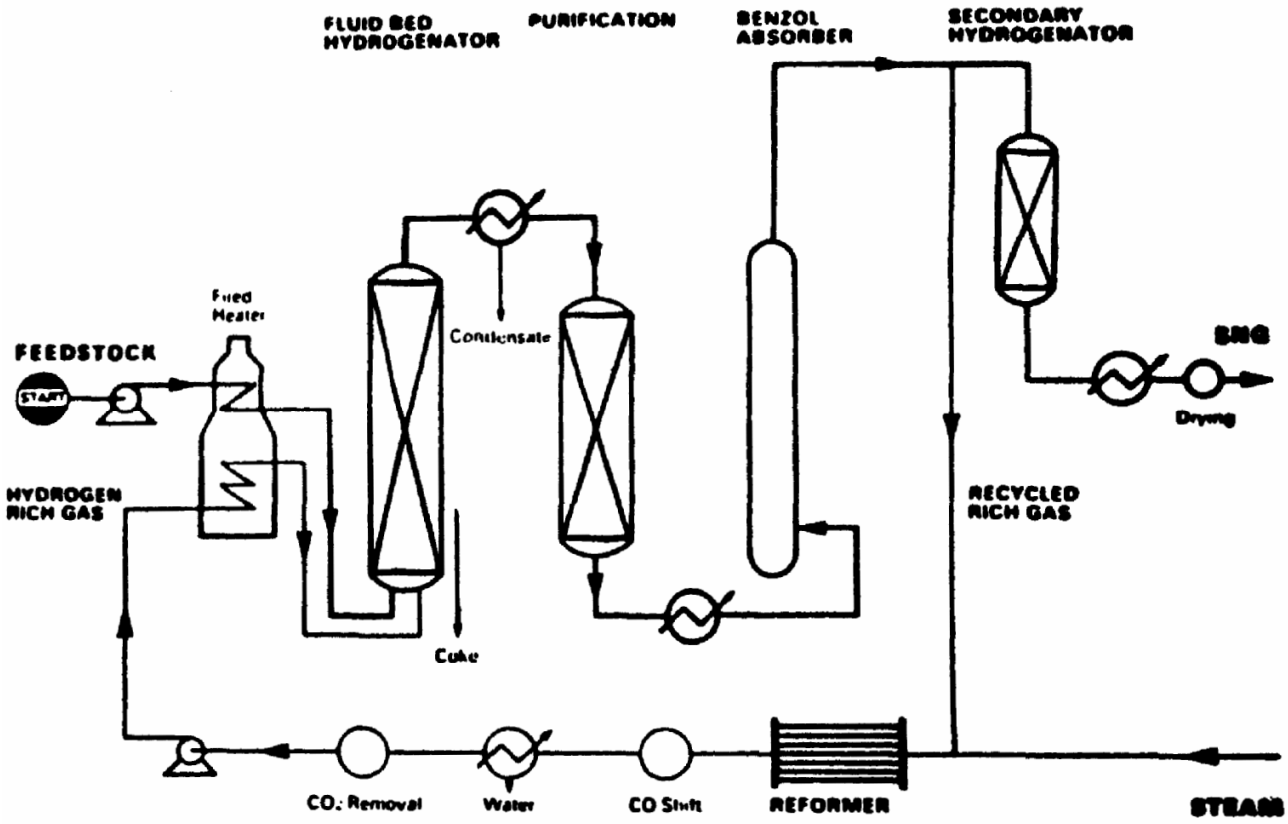

Figure 21.10: The Fluid-bed Hydrogenator Based SNG Process

21.7.4.3 Coal as Feed Stock

Because of the extensive reserves of coal, there has been a considerable interest in coal as a feed stock. The problem is that there have only been a few coal gasification plants built and these produce relatively low BTU gas.

The prime process that has been used is the 'Lurgi' process that has been developed in South Africa and used there for many years. There have been several other processes suggested and demonstration plants built using these processes. The prime difference in the processes is the means of gasifying the coal. This is done by burning part of the coal with O_2 to provide the heat for gasification, or by other external means such as in the CO_2 Acceptor process or the Hygas process.

The upgrading of the gas to 1000 BTUs requires considerably more process effort, particularly in the methanation section. Work on this part is being carried out in the demonstration plants. Also, methods of removing excess CO_2 are being reviewed.

a) Chemistry

All coal-to-gas conversion processes now under development are based on either the devolatilization of coal to produce a gas and a char using simple heating, or are based on a chemical approach in which the bulk of the carbon in the coal is chemically reacted to produce methane. The most promising processes utilize the chemical approach since devolatilization results in a char product as well as gas.

In the chemical approach, the overall reaction is:

Reaction 21-7

$$coal + water \rightarrow methane + carbon\ dioxide$$

Reaction 21-7, however, cannot be carried out directly, but must be achieved by combinations of the following reactions:

Reaction 21-8

$$Coal + H_2 \rightarrow nCH_4$$

Reaction 21-9

$$C + 2H_2 \rightarrow CH_4$$

Reaction 21-10

$$C + H_2O \rightarrow CO + H_2$$

Reaction 21-11

$$C + O_2 \rightarrow CO_2$$

Reaction 21-12

$$CO + H_2O \xrightarrow{Catalyst} CO_2 + H_2$$

Reaction 21-13

$$CO + 3H_2 \xrightarrow{Catalyst} CH_4 + H_2O$$

Reaction 21-14

$$CO_2 + 4H_2 \xrightarrow{Catalyst} CH_4 + 2H_2O$$

Each of the above seven reactions release heat.

The amount of each reaction utilized (Reaction 21-8 to Reaction 21-14) depends on the process under consideration and will, in general, dictate the overall efficiency of that process. Briefly, Reaction 21-8 is the devolatilization of coal in the presence of hydrogen which is used to maximize methane formation. Reaction 21-9 is the sought after hydrogasification reaction which produces methane directly and releases heat at the preferred point in the process where it can be utilized directly to assist in the gasification reactions. Reaction 21-10 is the work horse of the operation, since it produces the hydrogen needed to react with the coal. In all the processes, heat required is generated by Reaction 21-11.

In certain of the processes, Reaction 21-11 is conducted internally within the gasifier, while in others combustion occurs in another vessel or furnace and the heat or energy is transferred to the reactor. Because of chemical equilibrium limitations, the product gas emerging from a gasifier still has substantial amounts of hydrogen and carbon dioxides. After properly adjusting the hydrogen-to-carbon monoxide ratio to a value slightly in excess of 3 utilizing Reaction 21-12, and after removing the sulphur compounds and the bulk of the carbon dioxide, additional methane is produced by the catalyzed Reaction 21-13 and Reaction 21-14.

b) Processes

There are several processes proposed for the manufacture of substitute gas from coal. The main ones are discussed below:

Most processes have many common stages. The major difference is in the stage in which the coal is reacted. An overall coal processing sheet is shown in Figure 21.11.

The coal is initially mined, crushed, sized, then by means of heating, the volatile portion removed. The coal is then sent to hydrogasification, which can be one or more vessels and steps. The gas from this step is then sent to shift conversion for converting as much CO and H_2O into CO_2 and H_2 (Reaction 21-12). Next, the gas is purified and sent to a methanator (Reaction 21-13 and Reaction 21-14). It is finally dried and compressed as required.

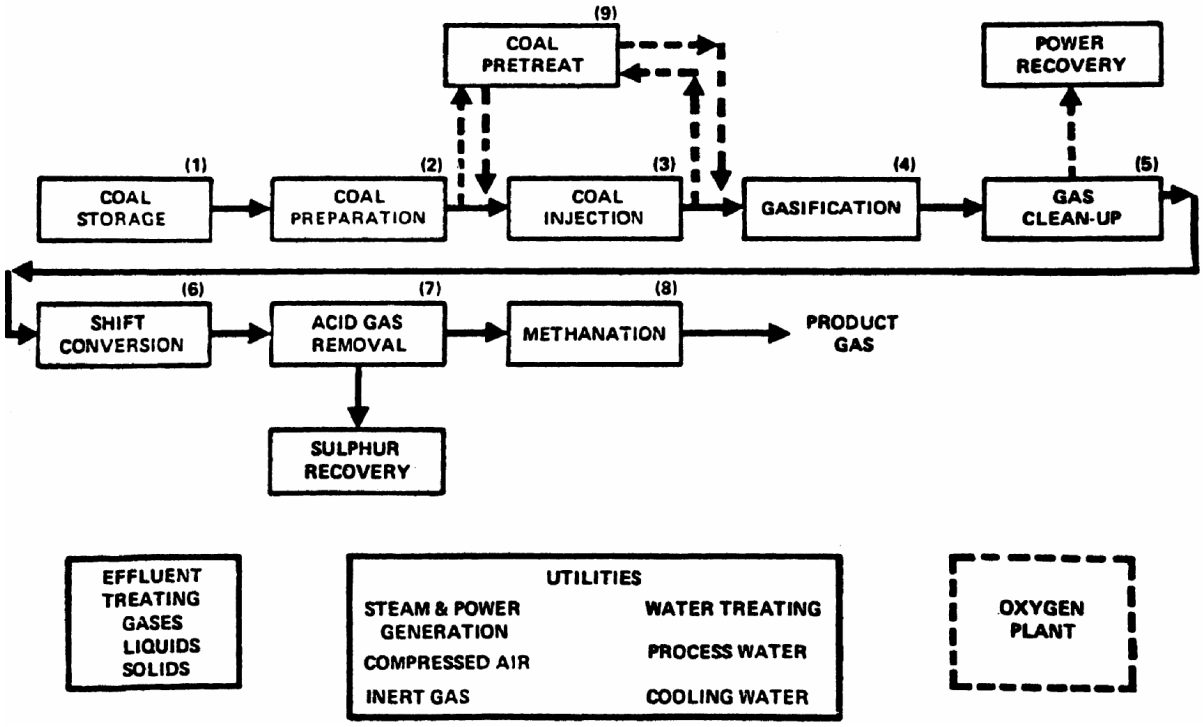

Figure 21.11: Coal Gasification Diagram

21.7.5 Coal Type Process

The type of coal available or planned to be used is quite important both in the choice of process and in the design of the gasifying equipment in the process selected. Figure 21.12 shows the relative content of moisture (H_2O), volatile matter and fixed carbon.

The moisture is generally removed in the pre-treating process. The volatile matter, mostly hydrocarbons, nitrogen and sulphur compounds, is gasified in the reactor and undesirable compounds removed later in the process (ie. H_2S, SO_2, NH_3, etc.).

A very important component to investigate is the ash content, which is a measure of the mineral content. This material will fuse between 1200ºC and 1800ºC and form slag. This slag has to be removed from the gasifier by special designs.

Another concern is the coke formation (rapid heating) qualities and the caking qualities (slow heating). These two phenomena can freeze an entire reactor into a single plastic mass of carbon. Generally, lignites have less tendency to glutinize.. Sometimes special pre-treatment can reduce the caking problem.

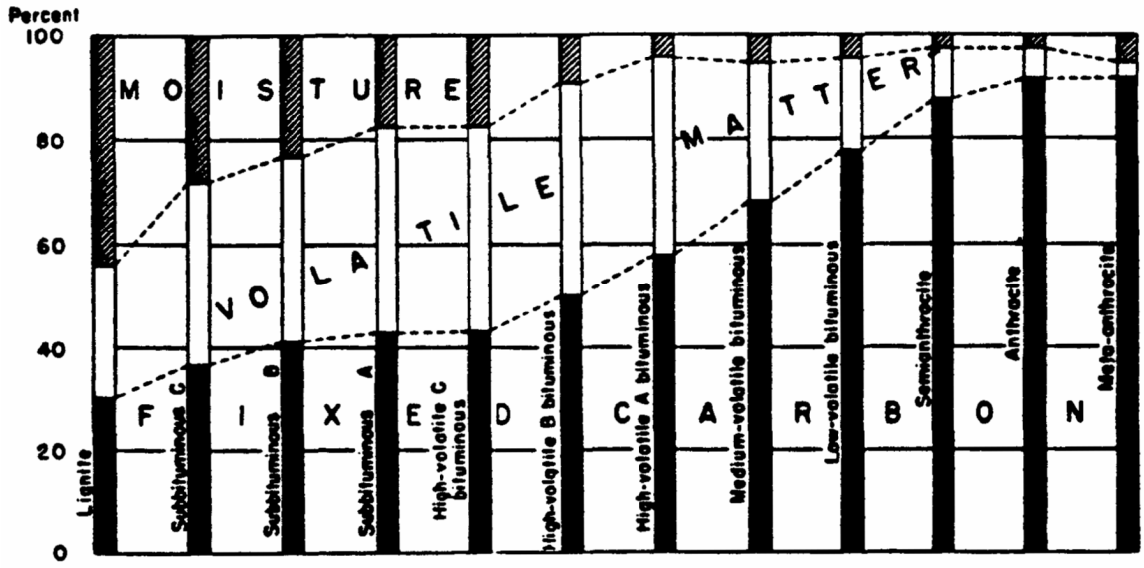

Figure 21.12: Coal Contents

21.7.6 Lurgi Process

This is the main process that has been developed commercially. A plant has been operating in South Africa using this process since 1936. This plant has been expanded several times. Also, plants are in operation in Scotland, Germany, and pilot plants in the United States. The heart of the process is the Lurgi gas generator which is shown in Figure 21.13. Various modifications to this gasifier are continuing to be made and are discussed by Shad and Hafke.

The process consists of the following steps:

1. Coal preparation - crushing, screening.
2. Coal gasification - a moving bed pressure gasifier.
3. Liquids recovery - separation of tars and other coal liquids.
4. Raw gas shift conversion.
5. Gas purification.
6. Methanation.
7. Drying and compression.

The coal received from the mine is first crushed and screened to provide a sized coal feed varying from 1/8 inch to 1-1/4 inches. The Lurgi gasifier operates at 300-500 psig. Coal is fed through a lock hopper system.

Steam and oxygen are introduced near the bottom of the gasifier and ash is removed via a lower lock hopper. A revolving grate supports the coal bed, governs the rate of ash discharge, and distributes the steam-oxygen mixture.

The Lurgi gasifier is in reality a moving-bed reactor. Synthesis gas (carbon monoxide and hydrogen) is generated in the bottom of the gasifier by the reaction of steam and devolatilized coal. Heat for this water gas reaction is provided by the reaction of oxygen and part of the coal. The hot synthesis gas passes up through the coal bed and effects devolatilization, drying, and

hydrogasification of the incoming coal. Bottom temperature is about 1,800ºF, and crude gas leaves the gasifier at about 1,100ºF.

The main reactions which occur are:

Reaction	Purpose	
$C + O_2 \rightarrow CO_2$	Provide heat.	Reaction 21-15
$C + H_2O \rightarrow CO + H_2$	Generate synthesis gas.	Reaction 21-16
$C + 2H_2 \rightarrow CH_4$	Methane production.	Reaction 21-17
Coal devolatilization	Gas and liquid production.	

The hot gas leaving the gasifier is cleaned of tar and dust in water quench vessels, steam generators, and coolers. Final temperature out of this section is about 360ºF.

The composition of the raw gas varies with the coal feed and gasifier operating conditions. The hydrogen to carbon monoxide ratio is important for the methanation step, and this ratio is adjusted in the shift converter by the following reaction:

Reaction 21-18

$$CO + H_2O \rightarrow CO_2 + H_2$$

The gas stream is then purified in a Rectisol plant which removes essentially all of the hydrogen sulfide, water, naphtha, organic sulphur compounds, gums and resins by absorption in cold methanol. Most of the carbon dioxide is also removed to obtain the right balance of carbon oxides and hydrogen to effect the desired methanation reactions. The gas stream must be free of sulphur compounds, naphtha, gums and resins prior to methanation so that the methanation catalyst will not be poisoned by them.

Methane is formed in the methanation process by the following reactions:

Reaction 21-19

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

Reaction 21-20

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$$

A simplified flow diagram of the Lurgi process is shown in Figure 21.14.

The coals that can be used in the Lurgi gasifiers are lignites, sub-bituminous and non-caking coals. As discussed earlier, caking coals agglomerate when heated and form a solid plastic mass, which makes the Lurgi gasifier inoperable.

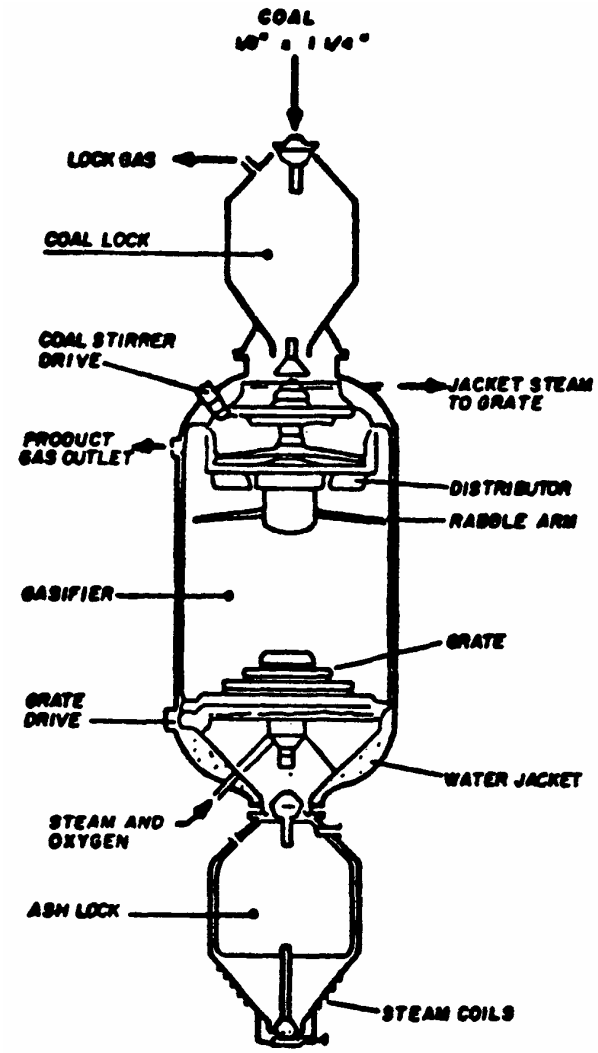

Figure 21.13: Lurgi Gasifier

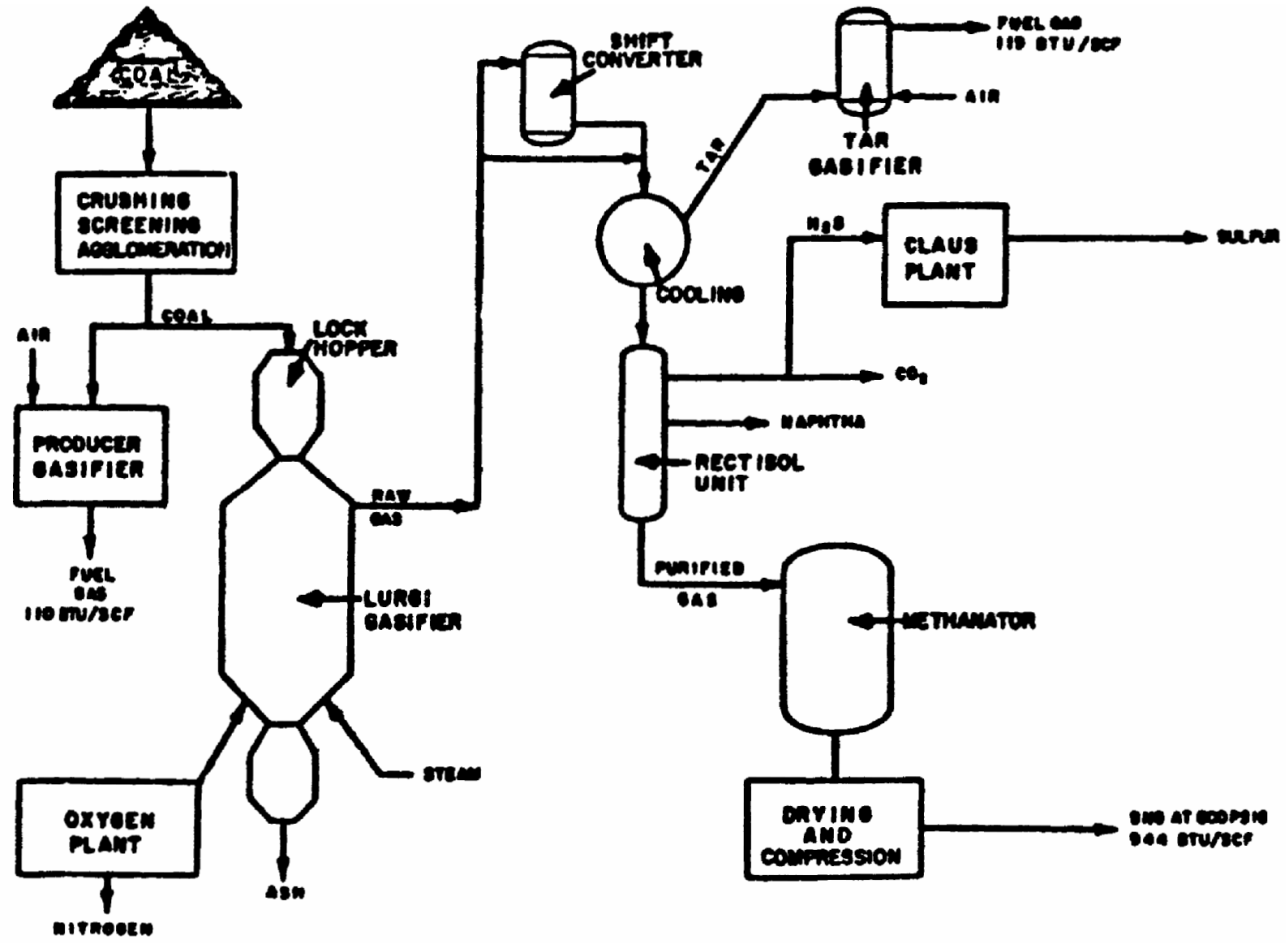

Figure 21.14: Simplified Flow Diagram of the Lurgi Process

21.7.7 CO_2 Acceptor Process

The CO_2 Acceptor Process is being developed by Consolidated Coal Company, a wholly owned subsidiary of Continental Oil Company, under the auspices of the Office of Coal Research. A 500 MCFD pilot plant has been completed and is being run at Rapid City, South Dakota. Lignites, with their high reactivity, are considered the best coal feedstocks for the CO_2 Acceptor Process.

A schematic of the gasification section of this process is shown in Figure 21.15. Hot, calcined dolomite (CaO), lignite, and steam are fed to the gasifier where several reactions occur:

1. The coal is devolatilized.
2. Synthesis gas is produced via the water gas reaction:

Reaction 21-21

$$C + H_2O \rightarrow CO + H_2$$

3. CO_2 is removed by reaction with CaO. This reaction also provides considerable exothermic heat.

Reaction 21-22

$$CaO + CO_2 \rightarrow CaCO_3$$

Char and dolomite are transferred to the regenerator where the remaining carbon is burned with air to provide the necessary heat to calcine the dolomite:

Reaction 21-23

$$CaCO_3 \rightarrow CO_2 + CaO$$

The gases produced in the gasifier require the same processing steps as previously discussed for the Lurgi Process except that the hydrogen to carbon monoxide ratio may not require adjustment. The action of the calcium oxide in removing CO_2 during the gasification step will result in significantly more hydrogen generation than in the Lurgi reactor.

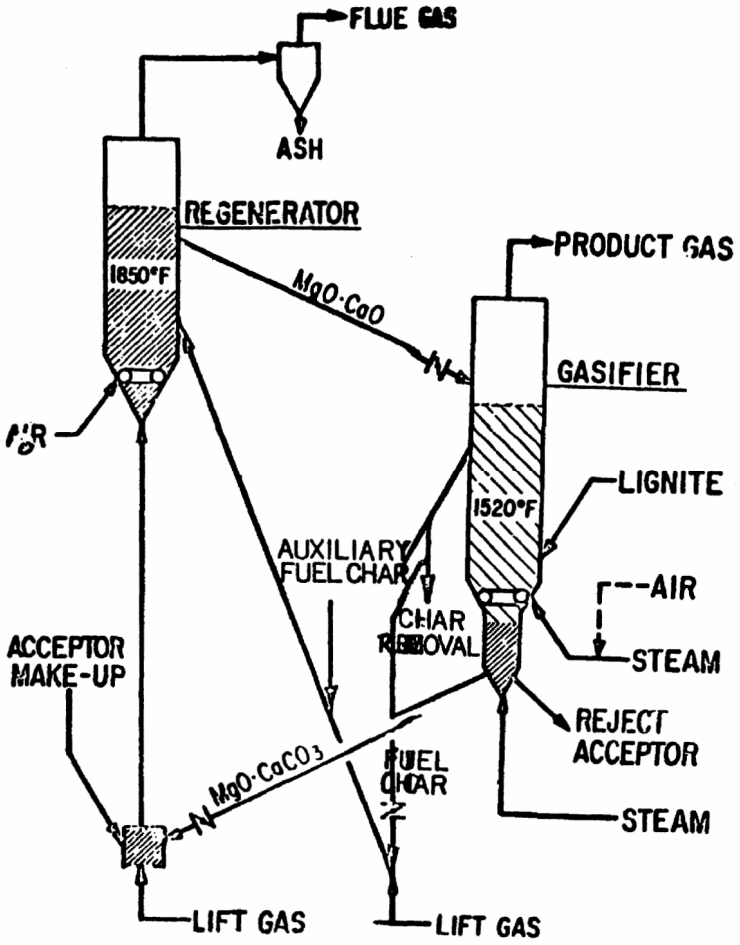

Figure 21.15: CO_2 Accept Process - Gasification Section

21.7.8 Hygas Process

A pilot plant running at 80T/day under the guidance of the Institute of Gas Technology in Chicago, Illinois, is presently studying another process scheme for gasifying coal. One version of this process is shown diagrammatically in Figure 21.16.

A slurry of coal in oil is pumped into the top section of a three-stage vessel operating at 1,000 to 1,500 psi where rising hot gases evaporate the carrier oil and heat the coal to about 1,500ºF. The coal falls into the first stage gasifier where it reacts with the hot gases from the second stage gasifier. The primary reaction is hydrogasification:

$$C + 2H_2 \rightarrow CH_4$$

and it is anticipated that about one-third of the methane in the final product will be formed in this section.

In the second stage gasifier, the partially converted coal reacts with incoming steam and hydrogen at about 1,800ºF. Synthesis gas and methane are formed in this section. The char is removed from the hydrogasifier and gasified with steam to provide the hydrogen-rich feed for this lower section.

IGT is considering alternative means of providing the heat required for converting the char to hydrogen by reaction with steam. The one shown in Figure 21.16 provides the heat of reaction electrically; another uses the oxygen and carbon reaction to provide heat; and a third uses the steam-iron process for producing hydrogen in an oxidation-reduction cycle.

The Hygas process yields a gas stream richer in methane than the raw Lurgi gas and should result in economics in the purification and methanation sections. Also, compression will not be required since the gasifier is designed to operate at 1,000 psi or higher. Similar downstream process steps, however, will be required for both the Lurgi and Hygas processes.

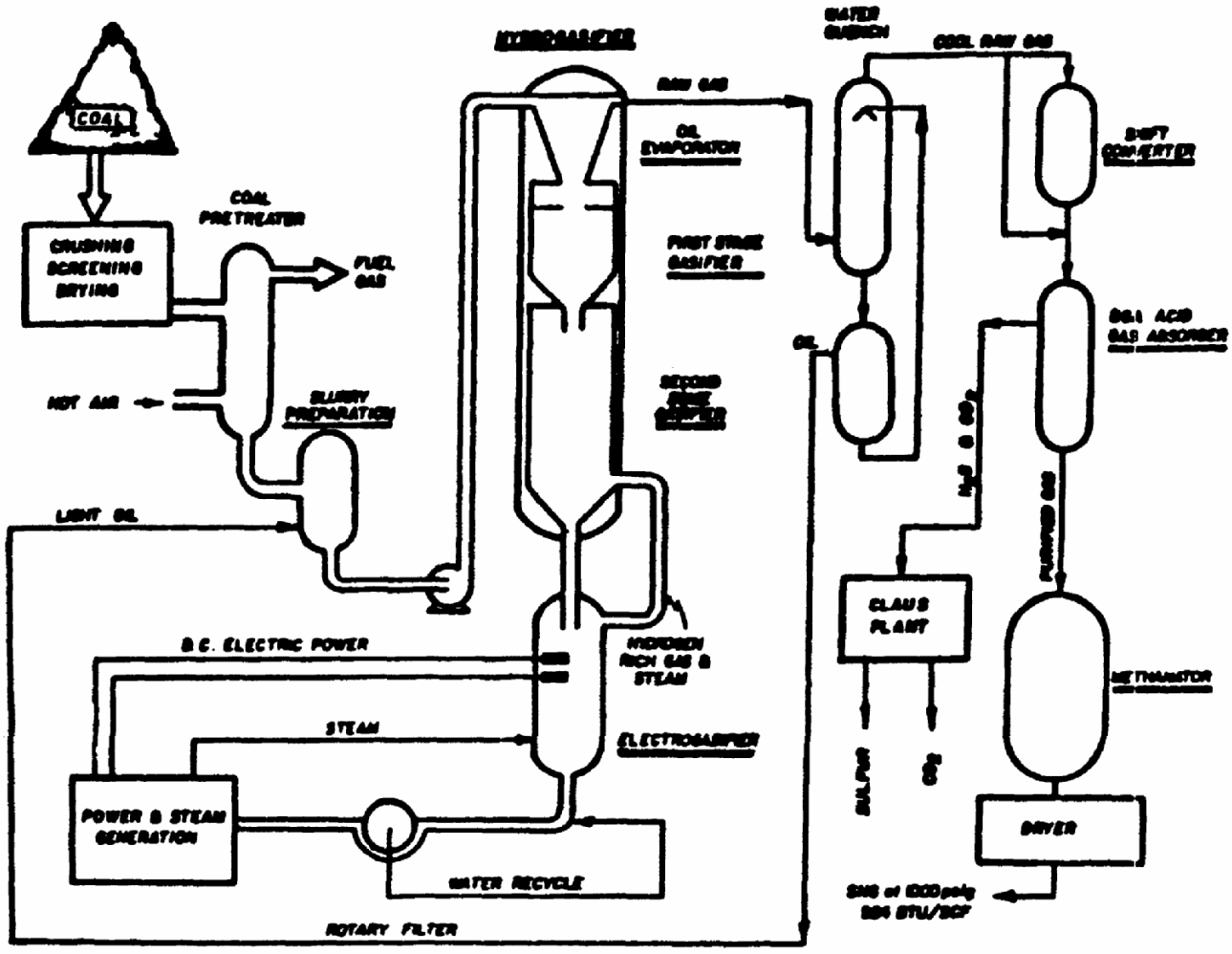

Figure 21.16: Flow Diagram of IGT Hygas Process

21.7.9 The Bi-Gas Process

A five ton/hour pilot plant using a process developed by the Bituminous Coal Research Inc. is under operation in Homer City, Pennsylvania.

The process consists of flash or rapid gasification of coal in an entrained bed using very hot gas from the slagging gasifier section in the bottom of the reaction vessel. These hot gases are produced by the reaction of oxygen with the partially gasified coal. Ash and slag are removed from the bottom of the reactor after being quenched with water.

21.7.10 Texaco Process

This is a variation of the Lurgi process in that the coal is fed to a reactor as slurry with water, mixed with oxygen and burned to provide heat for steam and gasification. The reactor operates at a very high temperature, which seems to eliminate such components as phenols, tars, and aromatic hydrocarbons. Also, because of the high temperature, the ash is produced as a molten liquid. The reactor itself, because of the high temperature, is ceramic lined. Reference "5" discusses the operation of a pilot plant in Germany that has been running since 1977.

21.7.11 Westinghouse Process

This process uses a fluidized bed gasifier using primarily recycle gas from the quench system as the fluidizing material for the coal. Steam and oxygen are added to the reactor and the normal gasifying reactions take place. Reference "6" discusses the status of the process.

There are other processes also being considered such as using sodium and potassium carbonate mixed with coal and sent to an embullating bed of catalyst, and combinations of the previously referred to processes.

21.8 General Comments

A very approximate number for the production of gas from coal is about 11 MCF/ton of sub-bituminous coal. Also, there is a requirement of about 200 gallons of water per ton of coal used. The investment costs for the processes are very high. The cost estimate for the Burham Z coal project, which is to produce 288 MMSCFD of 954 BTU pipeline gas, was $491,000,000 in 1973. This plant will use 23,000 tons of coal per day.

A good review of the various coal gasification processes is given by Furlong.

Also, a good review of the state of coal gasification is given by the Energy Resources Conservation Board of Alberta report on the subject.

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22.0 Environmental Considerations in Natural Gas Processing

- Consideration of flaring gas, discharges from sulphur plants, water treatment and disposal.

22.1 Introduction

This section of the course covers the handling of waste effluents that are emitted at various stages of production from the time the raw natural gas is recovered at the well to the processing plant where specification pipeline gas and other co-products are produced for sale in the commercial market. Operations such as the flaring of hydrocarbons, the discharge of effluent gases during the production of sulphur and the discharge of liquid wastes such as boiler blowdown, cooling tower waters and surface runoff from the plant site will be considered. Emissions will be divided into groups causing air pollution, water pollution, soil pollution and noise pollution. Acts and regulations related to the control of emissions in these categories will be discussed.

To grasp the magnitude of natural gas processing facilities contribution to pollution, here are some statistics on the emission of sulfur:

- The natural gas processing industry in Alberta emitted an estimated 375 000 long tons of sulphur in 1985.
- In the United States it is estimated that 30 000 000 long tons were emitted in 1971.

The serious effect of SO_2 emissions to the atmosphere may be appreciated by the result of the London smog of 1952 in which about 4,000 people died during the five day incident. The average SO_2 concentration was 1.3 ppm by volume.

Note: This section was originally written primarily by Dr. EL. Tollefson, Professor - Department of Chemical Engineering, The University of Calgary, prior to 1972. It was revised in July 1975 by Dr. D.M. Leahey, S.A. Zaidi, and E.M. Berlie of Western Research & Development Ltd., Calgary; further revisions have been made by Dr. EL. Toilefson and Dr. A.IL Younger, and most recently it has been revised from notes and lectures given by Dr. D.M. Leahey and D. McCoy of Canterra Energy. However as of 1992 this section is considerably out of date and needs extensive revisions.

22.2 LEGAL REGULATIONS (Legislative Controls)

Note: These apply to the Province of Alberta, Canada. For other provinces, states, etc. Check with appropriate governing bodies.

Five dominant pieces of legislation control developments and operations related to gas processing in Alberta. These are The Land Surface Conservation and Reclamation Act, The Clean Air Act, The Clean Water Act, The Water Resources Act and The Oil and Gas Conservation Act. Excepting the last named Act, all other Acts are administered by the Alberta Department of the Environment.

In addition, there are the regulations enacted pursuant to these Acts and the permits, guidelines and directives issued by the enforcing agency from time to time. However, the Energy Resources Conservation Board acts as the coordinator of the approval process and formal application is made

to them. They in turn, obtain approval of the proposal from the Alberta Department of the Environment.

When a site has been selected for a gas processing plant, the land must first be zoned industrial, if it is not so zoned, by the municipal agency passing the necessary bylaw. Application is then made to the regional planning commission, or other authority for a building permit.

Under the Land Surface and Reclamation Act, when a person plans any operation or activity that will result in surface disturbance, the Minister of the Environment may order a report on the assessment of the environmental impact, where it is considered in the public interest to do so. The order may ask for the report to contain an assessment of the impact of all or any of the following:

a) The conservation, management and utilization of natural resources.

b) The prevention and control of pollution of natural resources.

c) The prevention of noise and the control of noise levels resulting from the operation or activity in so far as they affect the environment in the vicinity of these operations or activities.

d) Economic factors that directly or indirectly affect the ability of the applicant to carry out measures that relate to the matters referred to in Clauses (a), (b) and (c).

e) The prevention of natural resources for their aesthetic value.

It should be noted here that those operations, not classed as gas processing, per se, which use gas or gas product as a raw material or fuel in the production or carbon black, ammonia, urea, ethylene, ethanol, methanol or any petrochemical products must apply to the Energy Resources Conservation Board for an Industrial Development Permit as provided for under Section 42 of the Oil and Gas Conservation Act.

The information to be submitted in support of the application includes the following: general description, background of owners and participants, process details, raw materials and fuel, upgrading of resources, marketing plans, capital cost and operating expenditures, employment, economic viability and financing; ownership and control policy, related industrial development, required government support, a brief on the environmental impact and the zoning authorization, and the overall impact on the Alberta economy and public interest.

With respect to the Clean Air Act, Clean Water Act and the Oil and Gas Conservation Act, effective April 18, 1973, the environmental management and pollution control responsibilities related to gas processing operations for the Department of the Environment and the Energy Resources Conservation Board were re-aligned and detailed in Board Information Letter IL-OG-72-20 and several subsequent directives. The following summarizes the changes:

1. Applications on environmental and energy resource aspects as filed with the Board for appropriate approvals by the Board and the Department.

2. Approvals, permits and licences issued by the Department shall be consolidated with others by the Board, and released to the applicant by the Board. This consolidated approval will prescribe conditions relating to both energy resources and environmental matters, including environmental control.

3. Inspection and surveillance of energy resource recovery, environmental conservation and pollution control features of gas plants normally are the responsibility of the Board, although, offsite monitoring checking will be carried out by the Department with corrective action directed by the Board.

4. Sour gas processing plants must prepare a monthly sulphur balance report as provided in the regulations under the Oil and Gas Conservation Act.

5. A monthly offsite air monitoring report must be filed with the Department along with a copy of the plant sulphur balance report.

Prior to the commencement of construction of a gas processing plant, an application for a permit to construct must be filed with the Board under the Oil and Gas Conservation Act. If the design of the plant has been finalized, this may also be the application for a license to operate. The application for approval to construct must cover such general matters as the location, conservation levels and pollution control features for the proposed scheme as detailed in Section 15.050(3) of the Regulations, including: (this act should be reviewed prior to making an application)

1. Maps of the area within a 10 mile (16 km) radius. Also including where practical, the location of all occupied buildings within 5 miles (8 km) of the plant.
2. A flow diagram along with a general narrative description of the scheme involved and the pollution control equipment provided, the maximum process capacity and the recover efficiencies for gas, natural gas liquids and sulphur.
3. Complete detail on feed composition and the overall plant material balance.
4. The maximum calculated ground-level concentration of air contaminants for the maximum design levels of their emission and discussion on methods for the control of smoke, sulphur dust and odour.
5. The design and size, and the composition of the effluent from the incinerator stack and the flare stack.
6. Methods and capacity of product storage.
7. Treatment and disposal methods for surface run-off water, process area run-off water, produced water with oil or gas, process waste water, sanitary sewage and other plant liquid and solid industrial waste.
8. Statements on the potential for accidents, emergencies and contingency plans to minimize damage, proposed methods for monitoring liquid and gaseous emissions, volume flow rate concentration of effluent pollutants, and ground-level concentration of gaseous pollutants.

Two other applications must accompany the submission filed with the Board under Section 38 of the Oil and Gas Conservation Act. These are the applications for a permit to construct under The Clean Air Act and The Clean Water Act. Although the data required on these three applications are duplicated, because of the joint responsibility and the organizational structure of the Department, it is considered advisable to prepare three different and distinct applications under each Act rather than attempt to prepare an integrated submission which will cover all aspects of the proposed scheme.

A copy of the complete application is forwarded to the Department by the Board, deficiencies are noted and corrected, and a public hearing may be held by the latter on environmental and resource considerations. Upon receipt from the Board of the Permit to Construct, these operations may commence at the site.

When the final design for the scheme is completed, the application for the license to operate under all three Acts is filed with the Board (see Section 15.050(4) Regulations Under Oil and Gas Conservation Act). This will include complete detail on generalized flow diagrams, material balance, recovery efficiencies, emissions of all types, emergency procedures and amended environmental impact assessment statements if an environmental impact statement was called for earlier. The latter may contain base line data collected since the application for the permit to construct was filed.

After the Department and the Board are satisfied as to the adequacy of all information, with the approval of the Minister of Environment (and the Lieutenant Governor in Council, if necessary) the License to Operate is issued.

Processing operations may then commence. The License to Operate will include specific statements respecting the location, raw gas feed rates, recovery efficiency required, maximum permissible emission levels of all pollutants, monitoring requirements, reporting requirements and other specific guidelines for the operator.

In the preparation of relevant applications and reporting with respect to gas processing operations, reference should also be made to the following guidelines, directives and standards:

Table 22.1: Energy Resources Conservation Board Letters, Guides and Reports

A. Energy Resources Conservation Board Letters, Guides and Reports	
Information Letter No.	**Subject**
ID OG-76-2	Emergency Procedure Plans for Sour-Gas Facilities.
ID 79-2	Minimum Distance Requirement Separating New Sour-Gas Facilities from Residential and Other Developments.
ID 80-2	Noise Control Guidelines.
IL 79-16	Revised Incinerator Stack Exit Temperature Criteria for Plants Processing Sour Gas.
IL 88-13	Sulphur Recovery Guidelines - Gas Processing Operations (also ERCB-AE 88-AA noted below)
IL 80-30	Application for Approval of Natural Gas Compressors.
Guides	
G-26	Sour Gas Processing - Plant Applications to the Energy Resources Conservation Board - A Guide to Content.
Reports	
ERCB 82-D	Sour Gas Processing in Alberta
ERCB-AE 88-AA	Sulphur Recovery Guidelines for Sour Gas Plants in Alberta
B. Alberta Environment	
Guides	Guide for Plume Dispersion Calculations (1980)**. Gas Processing Plants Waste Water Management Standards (1973) EIA Guidelines (1977)
Directives	
AMD 81-1	Oil and Gas Industry Air Monitors Directive (Note as of March 1988 there has been a new draft expanding this called AMD-86 covering all industry issued formal one not out yet).

** Models in this publication have been replaced by a new series. They are called "STACKS2, PLUMES2, SULDEP3, ADEPT and SEEC" and are available from Alberta Environment.

The standards of Air Quality for Alberta are given in Table 22.2: Alberta Air Quality Standards. Calculated and measured ground-level, or tree top level if applicable shall not exceed these concentrations.

Table 22.2: Alberta Air Quality Standards

Air Contaminant	Time Period	Concentrations Alberta Standard
Sulphur Dioxide		*ppm*
	Annual Arithmetic Mean	0.01
	Max. 24 hour concen.	0.06
	Max. 1 hour concen.	0.17

Air Contaminant	Time Period	Concentrations Alberta Standard
		ug/m^3
	Annual Arithmetic Mean	30
	Max. 24 hour concen.	150
	Max. 1 hour concen.	450
Suspended Particulates		*ug/m^3*
	Annual Geometric Mean	60
	Max. 24 hour concen.	100
Carbon Monoxide		*ppm*
	Max. 8 hour concen.	5
	Max; 1 hour concen.	13
		ug/m^3
	Max. 8 hour concen.	6
	Max. 1 hour concen.	15
Total Oxidants (as O_3)		*ppm*
	Max. 24 hour concen.	0.025
	Max. 1 hour concen.	0.08
		ug/m^3
	Max. 24 hour concen.	30
	Max. 1 hour concen.	100
Hydrocarbons		ppm
	Max. 3 hour concen.	--
		ug/m^3
	Max. 3 hour concen.	--
Fluorides (as HF)		ppm
	Max. 24 hour concen.	0.004 (Prov.Brd. of Health)
		ug/m^3
	Max. 24 hour concen.	3.3 (Prov.Brd. of Health)
Total Oxides of Nitrogen (as NO_2)		ppm
	Max. 1 year concen.	0.03
	Max. 24 hour concen.	0.106
	Max. 1 hour concen.	0.21
		ug/m^3
	Max. 1 year concen.	60
	Max. 24 hour concen.	200
	Max. 1 hour concen.	400
Hydrogen Sulphide		ppm
	Max. 24 hour concen.	0.003
	Max. 1 hour concen.	0.010
		ug/m^3

Air Contaminant	Time Period	Concentrations Alberta Standard
	Max. 24 hour concen.	4
	Max. 1 hour concen.	14
Soiling Index	90% of average readings per month less than Annual average less than.	coh
		1.0 (Prov. Brd. of Health)
		0.45 (Prov.Brd. of Health)
Total Dustfall		tons/sq.mi.
	30 days	15.0 res. area
	30 days	45.0 ind. & com.area
		mg/100 cm^2
	30 days	53 res. area
	30 days	158 ind. & com. area

Note: Alberta's standards are based on the ppm value and the ug/m^3 value was calculated from the base figure.

The Standards for Alberta Surface Water Quality are presented in Table 22.3: Surface Water Quality Criteria.

Table 22.3: Surface Water Quality Criteria

These criteria have been prepared in cooperation with the Provinces of Saskatchewan and Manitoba and represent water quality suitable for most uses either through direct use or prepared for use by an economically practical degree of treatment.

	Parameter	Criteria
1.	Bacteriology (Coliform Group)	1. In waters to be withdrawn for treatment and distribution as a potable supply or used for outdoor recreation other than direct contact, at least 90% of the samples (not less than five samples in any consecutive 30-day period) should have a total coliform density of less than 5,000 per 100 ml and a fecal coliform density of less than 1,000 per 100 ml. The Maximum Permissible Limit of total coliform organisms in a single sample shall be determined by the Provincial Board of Health based on the type and degree of pollution and other conditions existing within the watershed. 2. In waters used for direct contact recreation or vegetable crop irrigation the geometric mean of not less than 4 samples taken over not more than 30-day period should not exceed 1,000 per 100 ml total coliforms, nor 200 per 100 ml fecal coliforms, nor exceed these numbers in more than 20% of the samples examined during any month, nor exceed 2,400 per 100 ml total coliforms on any day.
2.	Dissolved Oxygen	A minimum of five mg/l at any time
3.	Biochemical Oxygen Demand (BODs)	Dependent on the assimilative capacity of the receiving water. The BODs must not exceed a limit which would create a dissolved oxygen content of less than five mg/l.
4.	Suspended Solids	Not to be increased by more than 10/mg/l over background value.
5.	pH	To be in the range of 6.5 to 8.5 pH units but not altered by more than 0.5 pH units from background value.
6.	Temperature	Not to be increased by more than 3ºC above ambient water temperature.
7.	Odour	The cold (20ºC) threshold odour number not to exceed eight.
8.	Colour	Not to be increased more than 30 colour units above natural value.
9.	Turbidity	Not to exceed more than 25 Jackson units over natural turbidity.
10.	Organic Chemicals	

	Parameter	Criteria
	Constituent	*Maximum Concentration (mg/l)*
	Carbon Chloroform Extract (CCE) (includes Carbon Alcohol Extract)	0.2
	Methyl Mercaptan	0.05
	Methylene Blue Active Substances	0.5
	Oil and Grease	substantially absent - no iridescent sheen
	Phenolics	0.005
	Resin Acids	0.1
11.	Pesticides	To provide reasonably safe concentrations of these materials in receiving waters an application shall not exceed 1/100 of the 48-hour $T1_m$. Persistent insecticides such as DDT, Aldrin, Dieldrin, Endrin, Heptachlor should not be used on or near surface waters.
12.	Inorganic Chemicals	
	Constituent	*Maximum Concentration (mg/l)*
	Boron	0.5
	Copper	0.02
	Fluoride	1.5
	Iron	0.3
	Manganese	0.05
	** Nitrogen (Total Inorganic and Organic)	1.0
	** Phosphorus as P04 (Total Inorganic and Organic)	0.15
	Sodium (as per cent of cations)	between 30 and 75
	Sulphide	0.05
	Zinc	0.05
	** These criteria are presently under study and may require adjusting according to naturally occurring concentrations or conditions.	
The predominant cations of sodium, calcium and magnesium and anions of sulphate, chloride and bicarbonate are too variable in the natural water quality state to attempt to define limits. Nevertheless, in order to prevent impairment of water quality, where effluents containing these ions are discharged to a water body the permissible concentration will be determined by the Provincial Board of Health in accordance with existing quality and use.		
13.	Toxic Chemicals	
	Constituent	*Maximum Concentration (mg/l)*
	Arsenic	0.01
	Barium	1.0
	Cadmium	0.01
	Chromium	0.05
	Cyanide	0.01
	Lead	0.05
	Mercury	0.0001
	Selenium	0.01
	Silver	0.05

	Parameter	Criteria
14.	Radioactivity	Gross Beta not to exceed 1,000 pCi/l Radium 226 not to exceed 3 pCi/l Strontium 90 not to exceed 10 pCi/l
15.	Unspecified Substances	Substances not specified herein should not exceed values which are considered to be deleterious for the most critical use as established by the Province Board of Health.

22.3 Air Pollution Control

22.3.1 Process Operations

Under no condition shall hydrogen sulphide bearing natural gas be released to the atmosphere. While inadvertent releases of hydrogen sulphide may take place, during operations or maintenance procedures, every effort must be made to immediately eliminate the source or ignite it. If the maximum permissible hourly average concentration of SO_2 or H_2S has been exceeded, or if there has been an inadvertent release of sour gas, appropriate reports must be made to the ERCB and the Department as required by the regulations and air monitoring directives. The effect of H_2S and SO_2 concentrations is shown in Table 22.4.

Table 22.4: Effects Of H_2S And SO_2 and Concentrations (No Particulate Matter)

	Hydrogen Sulphide	Sulphur Dioxide
Noticeable by odour.	0.02 - 0.1 ppm	3.0 ppm
Irritating to the eyes, nose or throat.	20 - 90 ppm	10.0 ppm
Accepted concentration for industrial working conditions.	10 ppm	5.0 ppm
Inhibits the human sense of smell.	100 ppm	--
Hazardous to life[*]	300 – 1,000 ppm	> 2,000 ppm
Produces instant death.	1% (10,000 ppm)	--
Threshold for plant damage.	20 – 40 ppm	0.3 ppm [**]
Tarnishes paint and silverware.	0.1 ppm	--
Accelerates rusting of iron and steel.	--	Varies with humidity, but SO_2 is worse than H_2S.

[*] At 300 ppm, the exposure may be over a matter of hours while at 1,000 ppm there is almost instantaneous unconsciousness.

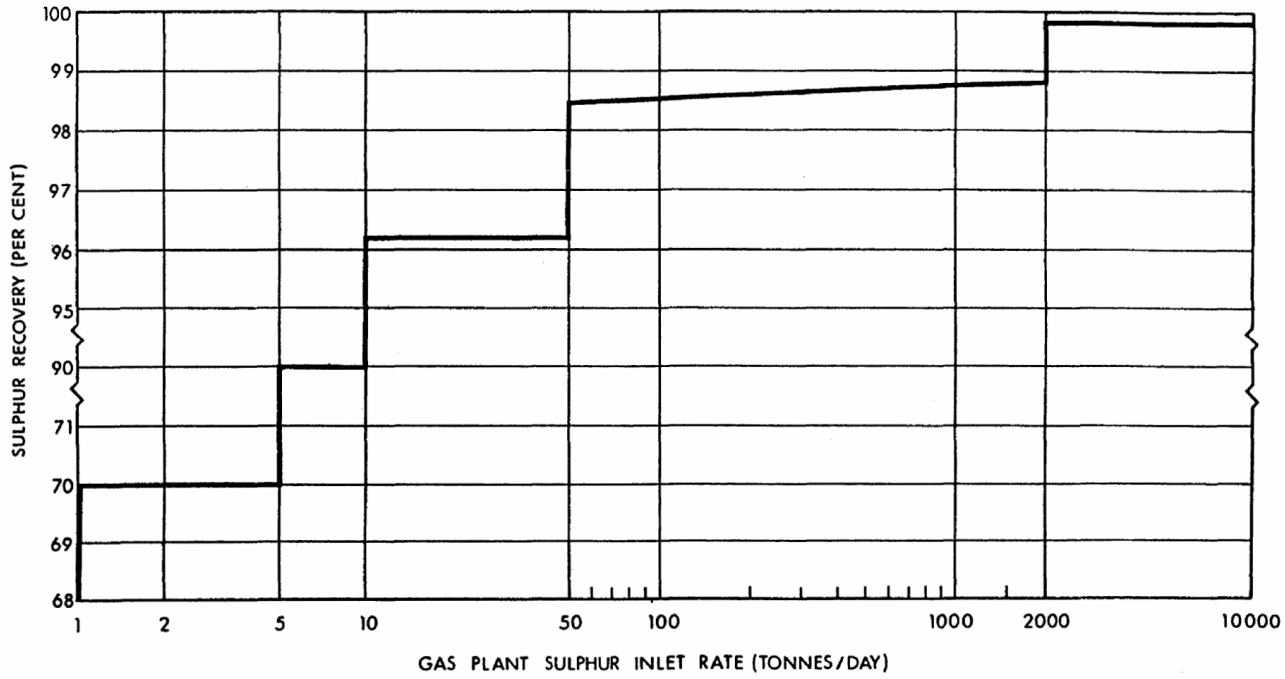
[**] The effect of sulphur dioxide varies with the type of vegetation, the time of exposure and the concentration during the exposure. For sensitive plants, 0.3 ppm for eight hours is the threshold condition and 1.0 ppm for one hour is similar.

The amount of sulphur dioxide that can be discharged to the atmosphere is a function of quarterly average sulphur recovery efficiency specified in the License to Operate issued by Alberta Environment. Guidelines for the emission of sulphur dioxide for new and expanding sulphur recovery plants were presented in the ERCB information letters No. IL-OG-80-24. These are summarized in Figure 22.1.

While the guideline as presented, suggested the process requirement that may be required to achieve the specified emissions, such as "3 stage Claus plant minimum" or "stack clean-up

required", these processes are not mandatory if they desired recovery efficiency can be achieved normally and on a quarterly basis through optimum design and well controlled operations.

At present consideration is being given to lower the inlet sulphur quantity at which sulphur must be removed and to the installation of zero emission plants in some cases.



Note: At normal operating conditions. Deduct 0.3% for quarterly average requirement.

Figure 22.1: Sulphur Recovery Requirements for Alberta Sour Gas Plants - August 1988

22.3.2 Design of Stacks

Gas processing plants must be designed such that treetop level hourly SO_2 concentrations resulting from their emissions do not exceed 0.17 ppm. There are six atmospheric situations which may give rise to air quality problems. They are: chimney downwash, building downdrafts, strong winds, intense thermal heating from the sun, plume trapping and irregular topography. A brief resume of the characteristics of each condition is given below.

1. *Chimney Downwash*: Downwash is produced by vorticies shed from the leeward side of a chimney. Stack effluent may be entrained into these vorticies, drawn down the side, and deflected to the ground. The downwash phenomenon usually occurs at high wind velocities. Problems associated with downwash may be avoided by having an effluent velocity in excess of the wind speed at all times.

2. *Building Downdrafts*: Effluent emitted from stacks may be entrained into highly turbulent wakes generated in the leeward side of buildings. Stacks located near buildings must be designed to avoid high ground-level pollutant concentrations that may occur as a result of this entrainment. Experience has shown that a turbulent wake generally extends to a height of about 1-1/2 times the building height. Stacks should therefore be designed so that they are at least 1-3/4 times as high as the surrounding buildings.

3. Strong Winds: The height to which a stack plume will rise above the ground decreases as wind speed increases. Ground-level concentrations of plume constituents tend to increase as the height of the plume above ground decreases. Therefore, strong winds may pose air quality problems. The Alberta Department of the Environment requires that all plant stacks be designed in a manner which ensures that the plume height above the ground is such that the hourly average treetop level concentration of SO_2 is always less than 0.17 ppm. The design techniques usually employ the Gaussian plume model assumption. This assumption essentially means that the plume is considered to have an oval cross section. The area of this cross section at any downwind distance may be obtained from curves supplied by Pasquill.

4. Intense Thermal HeatinQ From the Sun: During a hot summer day the sun heats the ground and produces thermals. These are characterized by regions of updrafts and downdrafts. During periods of light winds, thermals may be especially strong and well organized. If a plume is caught in a downdraft during these periods, it can be brought to the ground in a relatively undiluted state. Relatively low air quality can result.

   It is nearly impossible to counteract the so-called "looping plume" phenomena by building higher stacks. The problem can generally be solved only by a reduction in the stack emissions.

5. Plume Trapping: Plumes may become trapped between the ground and an elevated stable atmospheric layer. This layer acts as a lid preventing the upward dispersion of the plume. The trapped plume can only disperse downward towards the ground. Such conditions can impair ground-level air quality. Theoretical and observational studies show that for a given wind speed ground level pollutants concentration can be up to twice as great in this situation as compared to a normal situation. It is of interest to note that elaborate meteorological programs have been implemented in the United States with the purpose of controlling plant emissions during situations when plume trapping exists.

6. Terrain: Hills, river valleys, ridges, mountains and other topographic features have an influence on air flow and hence upon plume dispersion. Research into terrain induced problems not completed is very hard to generalize about these influences. It appears however, that terrain will usually affect plume dispersion in an adverse fashion. In most cases, terrain problems may be successfully resolved by proper stack design and plant location. Great care must be exercised in areas characterized by high winds and complex terrain.

Figure 22.2 shows some of the smoke plume patterns observed in the atmosphere.

As mentioned above, stacks must be designed for strong wind speed situations. For these design purposes, maximum treetop level SO_2 concentrations have been previously calculated by the Department of the Environment based on the Gaussian plume model:

Equation 22-1

$$X = \frac{Q.10^6}{\pi U \sigma_y \sigma_z} e^{-\left(\frac{H^2}{2\sigma_z{}^2}\right)}$$

where:

X = maximum ground-level value of the concentration (ppm by volume).

Q = rate of contaminate emission from a continuous point source, c.f.s. at 60ºF and 14.7 psia.

U = wind speed, f.p.s.

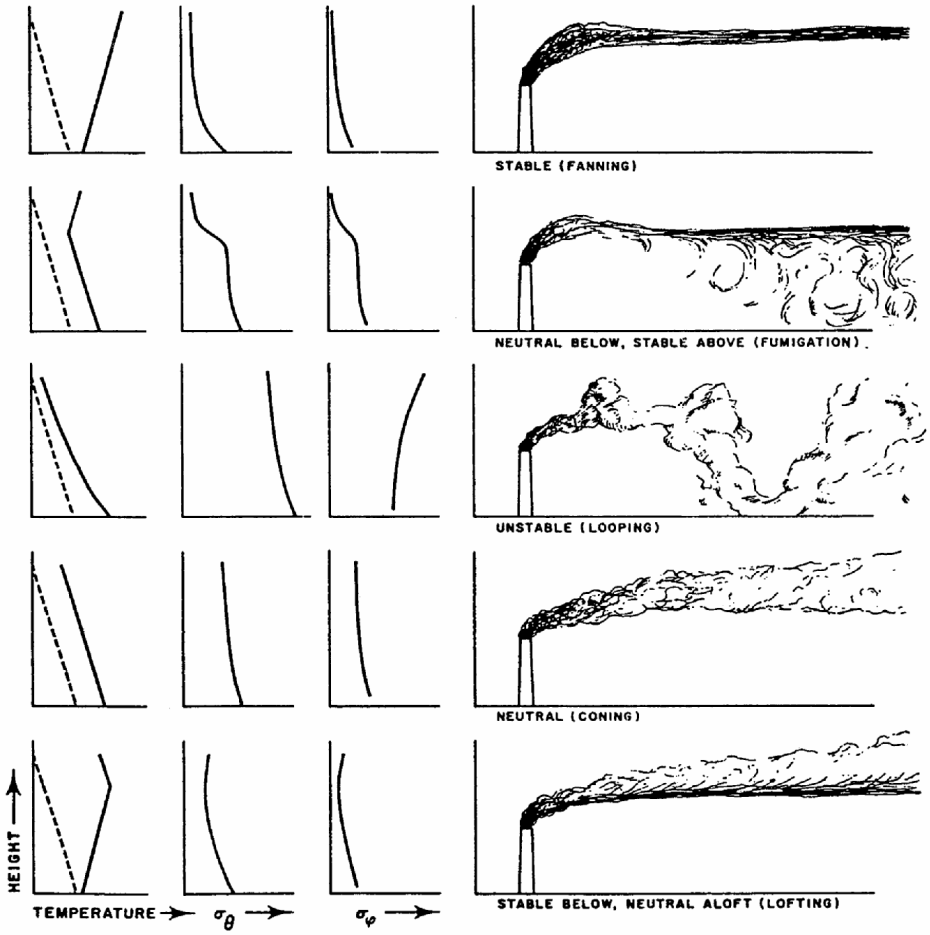
σ_y , σ_z = lateral and vertical standard deviations of the plume spread, respectively (ft).

H = effective height of the plume above the trees (ft).

Note: For an actual stack calculation please see "Guide for Plume Dispersion Calculations" by Alberta Environment (1980) with the programs as modified in 1988.

It should be stressed that the Gaussian plume model expressed in Equation 22-1 is for an average concentration over a period of greater than ten minutes. For such a average, the model gives a close approximation to actual conditions.

Problems arise in estimating ground-level concentrations from Equation 22-1 because of the need to determine the plume spread and the plume rise.



Note: Various types of smoke-plume patterns observed in the atmosphere. The dashed curves in the left-hand column of diagrams show the adiabatic lapse rate, and the solid lines are the observed profiles. The abscissas of the columns for the horizontal and vertical wind-direction standard deviations (σ_θ and σ_φ) represent a range of about 0º to 25º.

Figure 22.2: Atmospheric Transport And Diffusion Studies

The Department of the Environment presently uses values of plume spread as observed in England over flat terrain by Pasquill during neutral atmospheric conditions. Graphs of Pasquill's vertical and

lateral plume spreads for neutral conditions are given in Figure 22.3. Graphs for other cases are given in the literature.

The height of the plume above the ground is given by a consideration of the physical stack height, plume rise, plume dispersion, building location, terrain elevations and treetop height. An old formula that was used is as follows:

Equation 22-2

$$H = h_p + h_r - t - 3/4T$$

where:

h_p = physical stack height (ft).

h_r = plume rise due to bouyancy (ft).

T = elevation of the terrain above plant base (ft).

t = treetop height (ft).

Presently a computer program is used that takes more of the site condition into consideration, but the above formula can be used for a rough approximation.

Plume rise due to bouyancy is calculated by using the 2/3 law formula:

Equation 22-3

$$hr = \frac{C\,X^{\frac{2}{3}}\,F^{\frac{1}{3}}}{U}$$

where:

C = a dimensionless constant.

X = downwind distance (ft).

F = buoyancy flux ($ft^4\ s^{-3}$).

U = wind speed ($ft\ s^{-1}$).

Note: For an actual stack calculation please see "Guide for Plume Dispersion Calculations" by Alberta Environment (1980) with the programs as modified in 1988.

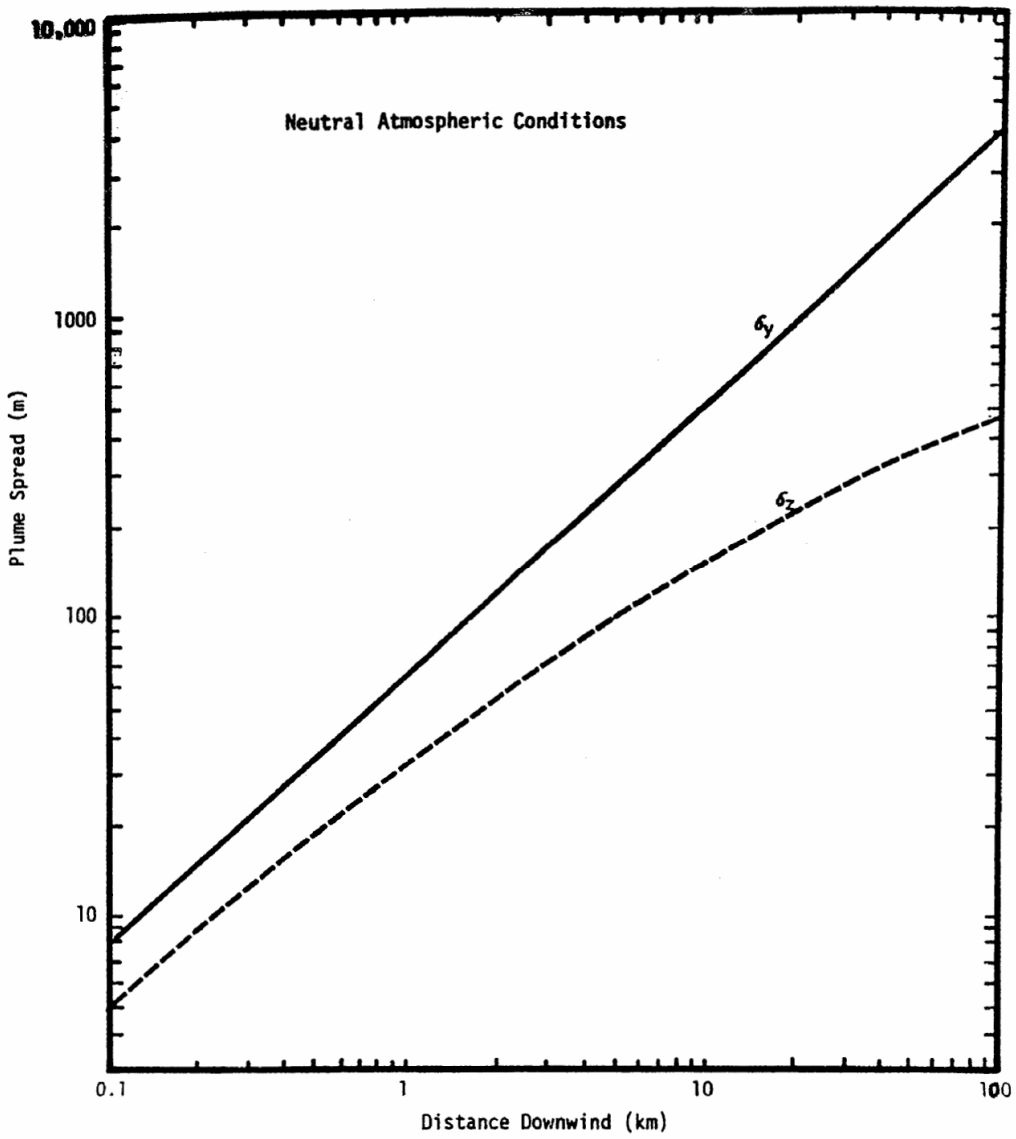


Figure 22.3: Vertical and Horizontal Plume Spread as a Function of Downwind Distances from the Source

For hot, dry effluents whose mean molecular weight is close to that of air, the buoyancy flux may be defined as:

Equation 22-4

$$F = \frac{g}{\pi}\frac{T_S - T_a}{T_a}Q_T$$

where:

g = acceleration due to gravity (ft s^{-2}).

T_s = absolute temperature of the stack gases, (ºR).

T_a = absolute temperature of the air, (ºR).

Q_T = rate at which total effluent is leaving stack ($ft^3\ s^{-1}$).

For sources of known heat release such as flare stacks, the buoyancy flux 'F' may be defined as:

Equation 22-5

$$F = \frac{g}{\pi} \frac{Q_H}{C_p \rho T_a}$$

where:

Q_H = rate of heat release (Btu s^{-1}).

C_p = specific heat of air at constant pressure (Btu $ºR^{-1}$ lb^{-1}).

ρ = density of dry air (lb ft^{-3}).

The above equation may be applied with any consistent set of units.

There have been many empirically derived values for the dimensionless constant C, ranging from 1.2 to 2.6. After reviewing the literature, Briggs recommended that a value of 1.6 be adapted. For conservative reasons, the Department of the Environment is considering using 1.2.

Physical stack heights for incinerators and flares are designed by ensuring that the treetop SO_2 concentrations calculated by using Equation 22-1 and Equation 22-2 are within acceptable limits. It should be stressed that the value used in Equation 22-1 for SO_2 emission (ie. Q) is normally 1.4 times the emission rate determined using the guideline quarterly average sulphur recovery efficiency at full plant load, as presented in ERCB information letter No. IL-OG-80-24. The actual emission multiplier will be considered for each application having regard for technology advances and proposed operating conditions.

22.3.3 Other Considerations

There are other requirements which must be considered in the design of incinerator and flare stacks. For incinerator stacks, these are as follows:

1. The stack gas oxygen concentration must be at least 1 per cent or such other higher value indicated to be necessary to ensure complete oxidation of the sulphur compounds in the plant tail gas.
2. The stack gas exit temperature must be at least 1,000ºF (538ºC).

Flare stacks are required at nearly all processing plants and at well installations when they are producing sour gas of an H_2S content greater than 1 per cent. The purpose of the flare stack is to burn vented combustible gases so that they can be discharged safely to the atmosphere. There are many important details in a flare stack such as the height, the diameter, burner tip design, the ignition system, the prevention of flashback and the prevention of smoke. This section covers only the first two items. The last item is covered later in this section. The other items are more of mechanical design details that are determined by stack manufacturers but certainly must be considered before putting in a flare stack. A good article, Flare Technology, is by J.F. Straitz.

The stack diameter can generally be determined by calculating the maximum quantity of flow expected up the stack at any one time. The maximum velocity allowed for this material is about 0.2 of the velocity of sound in the fluid being handled in plant flares. In flares at wells much higher velocities are often used, sometimes up to 0.8 the velocity of sound at maximum flow.

The height of the flare is generally determined by the controlling of two factors. One consideration is a flare stack flame out. In this case, the height should be such that a hazardous condition could not occur at ground level with concentrations of hydrocarbon above 2 per cent or 20,000 ppm.

Generally this is not a problem in a plant if the stack is at least the minimum height required under ERCB guidelines. The other consideration is to make the height so great that the heat intensity at the base of the stack is such that a man is not endangered when the value of 1,500 Btu/hr/ft^2 is set for the heat intensity. The height of the stack can then be found from the formula:

Equation 22-6

$$1500 = \frac{960W\sqrt{M}}{4\pi\left[H\left(H+120D\right)\right]}$$

where:

W = vapour being burned in lbs/hr.

M = molecular weight of vapour being burned.

D = stack diameter in feet.

H = stack height in feet.

Good references on flare stack design which go into more detail than the above discuss other factors by Kent, Hajek and Ludwig , Kannapell and Myers and Straitz. In the Kannapell and Myers article, the design of self-supporting stacks is discussed.

Other general requirements for flare stacks are:

1. That to prevent fires, the location of a flare stack should be approximately 500 feet from the nearest process equipment.
2. That at wells, a minimum stack height of 40 feet is required.

22.3.4 Regulating Smoke Production

The Alberta Clean Air regulation provides essentially smokeless flaring except on emergency condition. Smoke can be prevented by injection of steam or the use of excess air. This is discussed further on. Also Figure 22.4 gives the burning qualities of hydrocarbons.

Section 603 of the Alberta Energy Resources Conservation Board regulations state that no oil, gas or oily waste shall be burned except under conditions of controlled combustion where there is no significant or visible emission of smoke. Permission is required if burning is likely to produce smoke. Production of smoke due to burning of a well or plant effluent under emergency conditions must be reported to the Board immediately.

A knowledge of the burning qualities of hydrocarbons is useful in deciding to what degree they are likely to cause smoke production. The burning qualities of various hydrocarbons are shown in Figure 22.4. It will be observed that paraffinic hydrocarbons produce relatively little smoke. Olefins, especially the C_3 and C_4 homologues, are prone to produce smoke. Alkynes, diolefins and aromatics, show a much greater tendency to produce smoke. With proper burner design and air-to-fuel ratio, any of these materials can be burned without smoke production. Conditions for adequate mixing of the fuel and air at the proper time with a sufficiently high ignition temperature and a sufficient volume to allow complete combustion may be combined to produce a smokeless flame.

22.3.5 Elimination of Smoke in Flaring Waste Gases

Smoke may be eliminated from flaring waste gases either by steam or air injection.

22.3.5.1 Steam Injection

Steam may be added to an air-waste fuel mixture as it is released from a flare stack in order to suppress smoke. Use of steam increases mixing by causing turbulence. In addition, air is aspirated into the flame. The steam itself reacts chemically with cafbon particles, thus yielding oxygenated products which burn invisibly.

Reaction 22-1

$$C + H_2O \rightarrow CO + H_2$$

Reaction 22-2

$$C + H_2O \rightarrow CO_2 + H_2$$

Use of steam also prevents polymerization of some intermediate products reducing the tendency for aerosol formation. The effect of addition of steam in the reduction of smoke from a waste gas containing unsaturates is shown in Table 22.5.

22.3.5.2 Air Injection

Rather than use steam, some operators have tried supplying air at the edge of pits during burning operations. While the technique is effective in reducing smoke, the need for large volumes of low pressure air at the site creates problems.

An effective smokeless flare is one developed by John Zink. This burner consists of concentric pipes with hydrocarbon in one and air with a blower in the other. The air flow is boostered by a blower that controls the air flow to the stack. Other smokeless and noiseless flares have been developed by National Airoil Burner Co. Inc. An excellent state-of-the-art paper on smokeless flares has been presented by Vanderlinde.

Table 22.5: Use of Steam for Suppression of Smoke

Percent by Weight Unsaturates in Gas	Pounds of Steam Per Pound of Gas	
	No Smoke	*No Trailing Smoke*
0	0.15	0.12
5	0.28	0.21
10	0.41	0.30
15	0.55	0.40
20	0.68	0.49
25	0.81	0.59
30	0.95	0.68

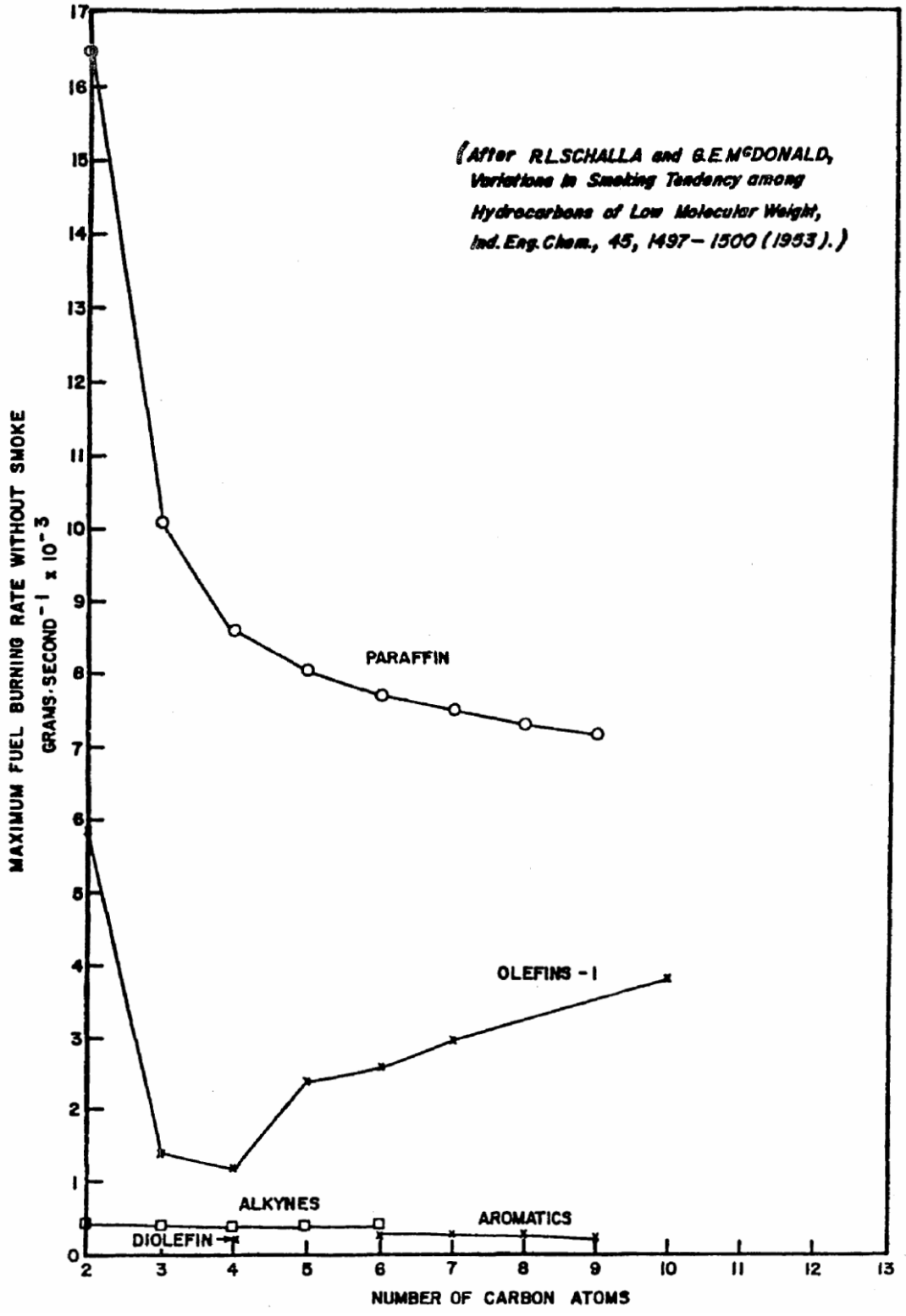


Figure 22.4: Burning Qualities of Hydrocarbons

22.3.6 NOx Concentrations in the Atmosphere

There is now a standard for the amount of NOx that may be discharged from various types of combustion processes. Figure 22.5 shows the concentration of NOx versus stack height and brake HP of compressor.

For other pieces of equipment it is generally necessary to refer to the manufacturer of it.

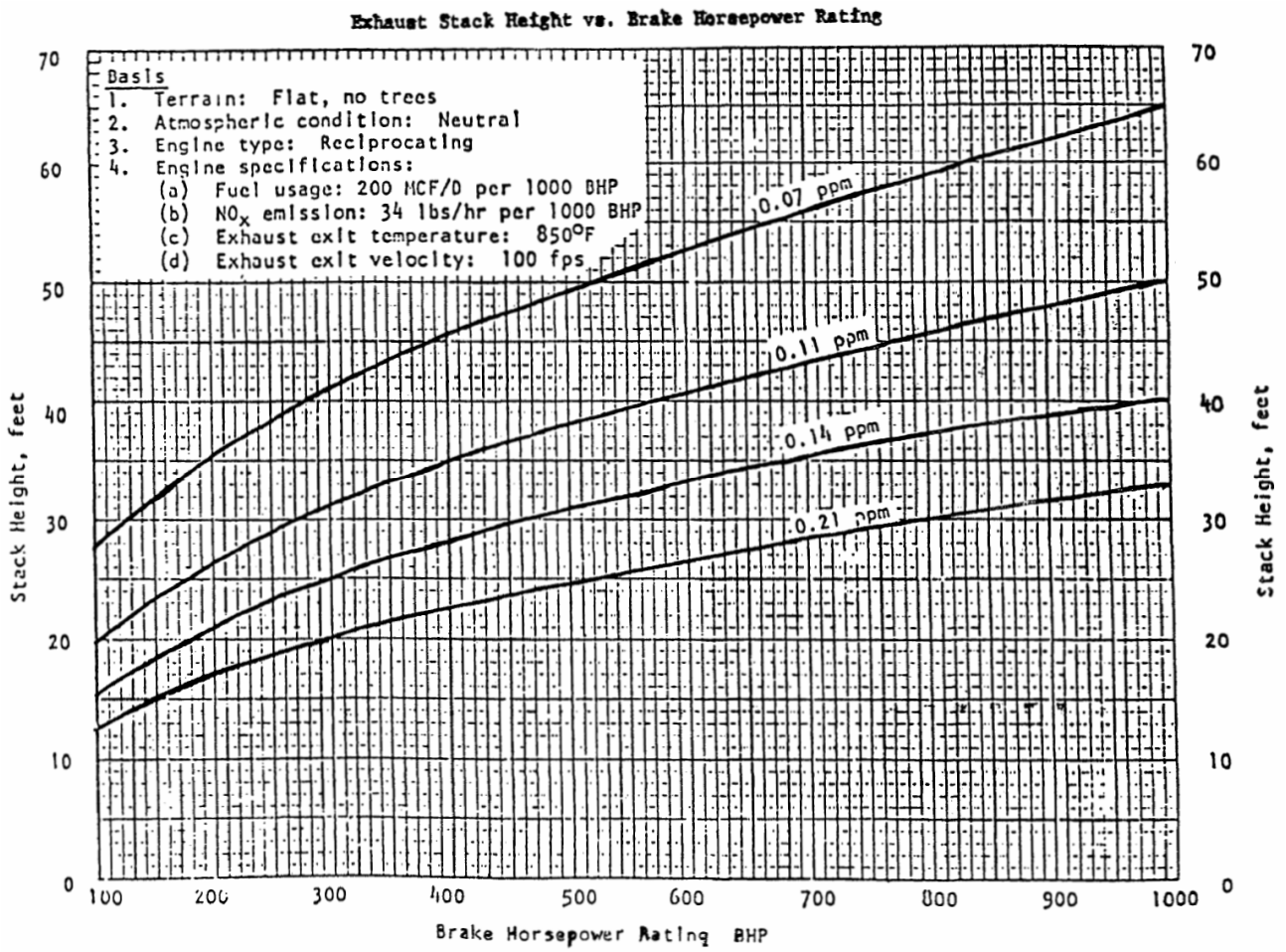


Figure 22.5: Natural Gas Fired Compression Facility

22.4 Treatment of Aqueous Wastes from Gas Plants

The aqueous wastes from gas processing plants contain a diverse range of pollutants including oil, hydrogen sulphide, dissolved solids, B.O.D. and C.O.D. - producing chemicals, and toxic metal ions. The actual amount of aqueous wastes originating from a specific plant would be qualitatively and quantitatively related to the size of the plant and to the number of unit operations employed (ie. absorption, dehydration, compression, refrigeration, fractional distillation, cooling, etc.). It is possible, therefore, to predict the type of pollutants that would be present in the aqueous wastes produced by a plant.

Figure 22.6 shows a simplified flow sheet for a typical sour gas treatment plant and indicates where various aqueous waste streams originate. The effluent streams can be broadly characterized as one or more of the following types:

a) oily.

b) high in sulphides.

c) high in dissolved solids.

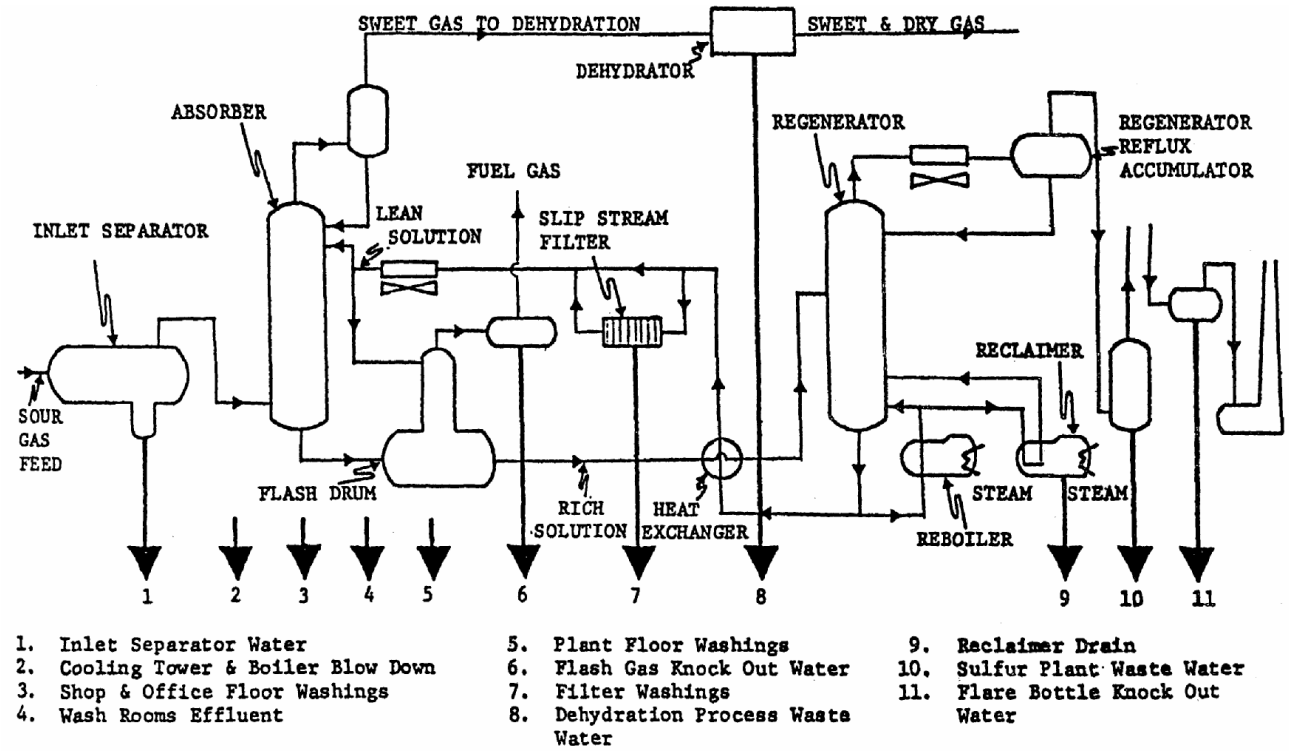
d) high in suspended solids.

e) high in dissolved organics.

f) high in metal ions.

Based upon these characteristics, a nomenclature system can be defined to classify the streams according to the type of pollutants that could be present in significant amounts. Table 22.6 lists the pollutants that can be expected in some of the source streams, and gives the estimated flow rates for these streams.

The proper segregation of effluent streams is extremely important in the overall design of effluent handling, treatment and disposal systems. The segregation scheme can vary and would depend on, the effluent treatment requirements, and the disposal system proposed for the effluent.



1. Inlet Separator Water
2. Cooling Tower & Boiler Blow Down
3. Shop & Office Floor Washings
4. Wash Rooms Effluent
5. Plant Floor Washings
6. Flash Gas Knock Out Water
7. Filter Washings
8. Dehydration Process Waste Water
9. Reclaimer Drain
10. Sulfur Plant Waste Water
11. Flare Bottle Knock Out Water

Figure 22.6: Flow Sheet for a Typical Sour Gas Plant and its Various Aqueous Waste Streams

Table 22.6: Estimated Flow Rates of Some Aqueous Waste Streams for a Hypothetical* Sour Gas Plant

No.	Waste Stream Description	Flow Rate (U.S. Gallons/Day)	Nature of Major Pollutants
1.	Inlet Separator Water.	6,500	Hydrogen Sulphide, Hydrocarbons, Hydrate inhibiting chemicals, possibly salt.
2.	Cooling Tower and Boiler Blow Down.	2,000	Dissolved Solids, Alkalinity, Phosphates, Chromates, Other Chemicals.
3.	Plants and office floor washings.	1,500	Hydrocarbons, Oil and grease, sweetening solution, Suspended solids.
4.	Washrooms effluent.	500	Biodegradable organics, Coli and bacteria.
5.	Filter washings.	2,800	Sweetening solution.
6.	Dehydration process water.	2,000	Hydrocarbons, Dehydrating solution.
7.	Reclaimer Drain.	60	Degradation products, Sweetening solution.
8.	Sulphur plant knock out water.	800	Hydrogen sulphide, Hydrocarbons, Sweetening solution.

* 150 MMSCFD Sour Gas Feed Containing 5% H_2S

22.4.1 Effluent Treatment Requirements

Gas processing plants in Alberta are subject to the Clean Water Act of the Province. The Department of the Environment, in cooperation with the Energy Resources Conservation Board, prepares and implements the regulations concerning the release of water contaminants from these plants. To minimize the release of pollutants, waste water quality standards are revised from time to time in the light of new developments in the field of water pollution control.

Currently, the gas processing plants in the province are required to comply with the waste water quality standards as set forth in the "Gas Processing Plant Waste Water Management Standards", published in September 1973, by the Department of the Environment. Explanations and interpretations of various terms used in the above mentioned document are contained in the "Guide- lines for the Gas Processing Plant Waste Water Management Standards", published by the Department in November 1973. For the purpose of implementing water quality standards, the gas processing plants in Alberta have been divided into three categories: A, B, and C. Table 22.7 summarizes the required waste water management programs for the three categories of gas plants. The applicable effluent quality standards are given in Table 22.8 through Table 22.10.

The effluent quality standards clearly reflect the need for and importance of segregating the major waste water streams (ie. surface runoff, process wastes, sanitary sewage, etc.) and disposing of them by suitable methods.

22.4.2 Disposal Methods

The segregation system for process waste water streams would depend upon the proposed ultimate disposal method for the streams.

The process aqueous wastes from the gas processing plants can be disposed of by:

- Subsurface injection.
- Evaporation.
- Release to surface streams.

Figure 22.7 through Figure 22.10 show four different segregation disposal systems. Depending upon the ultimate disposal methods chosen, one of these should satisfy the need for the disposal of the aqueous wastes from a gas processing plant. The three disposal methods mentioned earlier are discussed in the following paragraphs:

22.4.2.1 Subsurface Injection

Subsurface injection of process waste waters from gas plants is currently considered by the Alberta Department of Environment to be an acceptable disposal method and is being used increasingly in the province. Nearly all the sour gas plants in Alberta inject at least a part of their aqueous wastes into subsurface strata.

The concept of injection of liquids into underground space is not new. The oil industry has been using brine disposal wells for more than half a century. This technique has become increasingly attractive in recent years as a result of more stringent requirements for surface water pollution control.

In fact, the use of such terms as "treatment" and "disposal" for deep-well injection of industrial wastes is misleading. It is just a method of long term storage, usually, pre-treated waste in underground space. A number of authors have urged caution in the utilization of deep-wells for industrial waste water storage (12), (13) and (14). Some of the basic arguments against deep-well injection of industrial waste, as listed by Toffelmire et al, are:

1. There is always a potential risk of fresh water contamination through well failure, geological faults, or unknown abandoned wells.
2. The long-term geological effects of the injection and location of waste are not well known.
3. Any adverse effect caused by injection will be difficult to correct.

Milne has discussed the deep-well injection system for sour gas plant waste waters in detail. He proposed the following treatment of aqueous wastes before deep-well injection (assuming that the waste water would be compatible with the formation):

1. Gravity settling in a storage basin to remove large settleable particulate matter.
2. Filtering through diatomaceous earth to remove nonsettleable particulate matter.
3. Sterilization, by heating to 240ºF for 15 minutes, to kill bacteria.

Walter and Stewart have described some simple laboratory tests which may be used to investigate the compatibility of the waste waters with the formation. Various authors, (18), (19) and (20) have described the well construction for liquid waste injection. Figure 22.11 and Figure 22.12 show the details of an injection well and the pre-treatment facilities for a typical disposal scheme.

Table 22.7: Schematic of Waste Water Management Programs

Classification	Waste Water Type	Declaration	Required Action or Treatment	Ultimate Disposal
Class A Plant (no sulphur recovery - no discharge to watershed).	Surface run-off.	nominal average design flow rate of surface run-off.	control and diversion.	discharge to surface water body.
	Plant waste water.	no discharge.	--	evaporation, injection or alternative method but not to surface body or ground water.
	Sanitary waste water.	--	minimum 90% treatment.	discharge to surface water body.

Class B Plant (Recovers sulphur - no discharge to watershed).	Surface run-off.	nominal average design flow rate of surface run-off.	control, diversion, treatment and monitoring to Table 8 standards.	discharge to surface water body.
	Plant waste water.	no discharge.	--	evaporation, Injection or alternative method but not to surface body or ground water.
	Sanitary waste water.	--	minimum 90% treatment.	discharge to surface water body.
Class C Plant (a sweet or sour gas plant; discharges plant waste to watershed on a continuous basis).	Surface run-off.	nominal average design flow rate of surface run-off.	control, diversion, treatment and monitoring to Table 8 standards.	discharge to surface water body.
	Plant waste water.	nominal average flow rate of plant waste water.	control, treatment and monitoring to Table 7 standards and concentration variations.	discharge to surface water body.
	Sanitary waste water.	--	minimum 90% treatment.	discharge to surface water body.

Table 22.8: Standards of Treated Process Waste Waters Released from Class C Gas Plants to Watershed Areas

Column 1	Column 2	Column 3
Water Contaminant or Constituent	**Max. Allowable Emission Rate or Value**	**Max. Allowable Concentration or Value**
Chemical Oxygen Demand	1000 lbs/BCF *	500 mgm/l
Suspended Solids (Total)	100 lbs/BCF	50 mgm/l
Sulphide (Total)	1 lbs/BCF	0.2 mgm/l
Ammonia Nitrogen	50 lbs/BCF	10 mgm/l
Oil and Grease	100 lbs/BCF	10 mgm/l
Threshold Odour Number	100 units	100 units
pH	6.5 - 9.5 units	6.5 - 9.5 units

* lbs/BCF means pounds per one billion cubic feet of raw gas charged to process units based on the reference gas charge rate.

Table 22.9: Standards of Treated Surface Run-off Water Released from the Developed Area of Class B and C Gas Plants to Watershed Areas

Column 1	Column 2	Column 3
Water Contaminant or Constituent	**Max. Allowable Emission* Rate or Value**	**Max. Allowable Concentration or Value**
Chemical Oxygen Demand	50 lbs/day	50 mgm/l
Suspended Solids (Total)	25 lbs/BCF	25 mgm/l
Sulphide (Total)	0.1 lbs/day	0.1 mgm/l
Ammonia Nitrogen	5 lbs/BCF	5 mgm/l
Oil and Grease	5 lbs/BCF	5 mgm/l
Threshold Odour Number	50 units	50 units
pH	6.5 - 9.5 units	6.5 - 9.5 units

* Based on 100,000 gallons per day discharge.

Note: The emission rate of release shall be prorated if the design flow rate of surface run-off varies from 100,000 gallons per day.

Table 22.10: Standard of Toxic Elements

*Toxic Elements (Total)	Concentration (mgm/l)
Arsenic	0.25
Barium	1.0
Cadmium	0.10
Chromium (hexavalent)	0.30
Copper	1.0
Cyanide	0.025
Iron	1.0
Lead	0.10
Manganese	1.0
Mercury	0.0005
Nickel	1.0
Selenium	1.0
Silver	1.0
Vanadium	0.10
Zinc	1.0

* The aggregate total toxic element limit of any individual liquid effluent discharged from a Class C gas plant shall not exceed 5 mg/l.

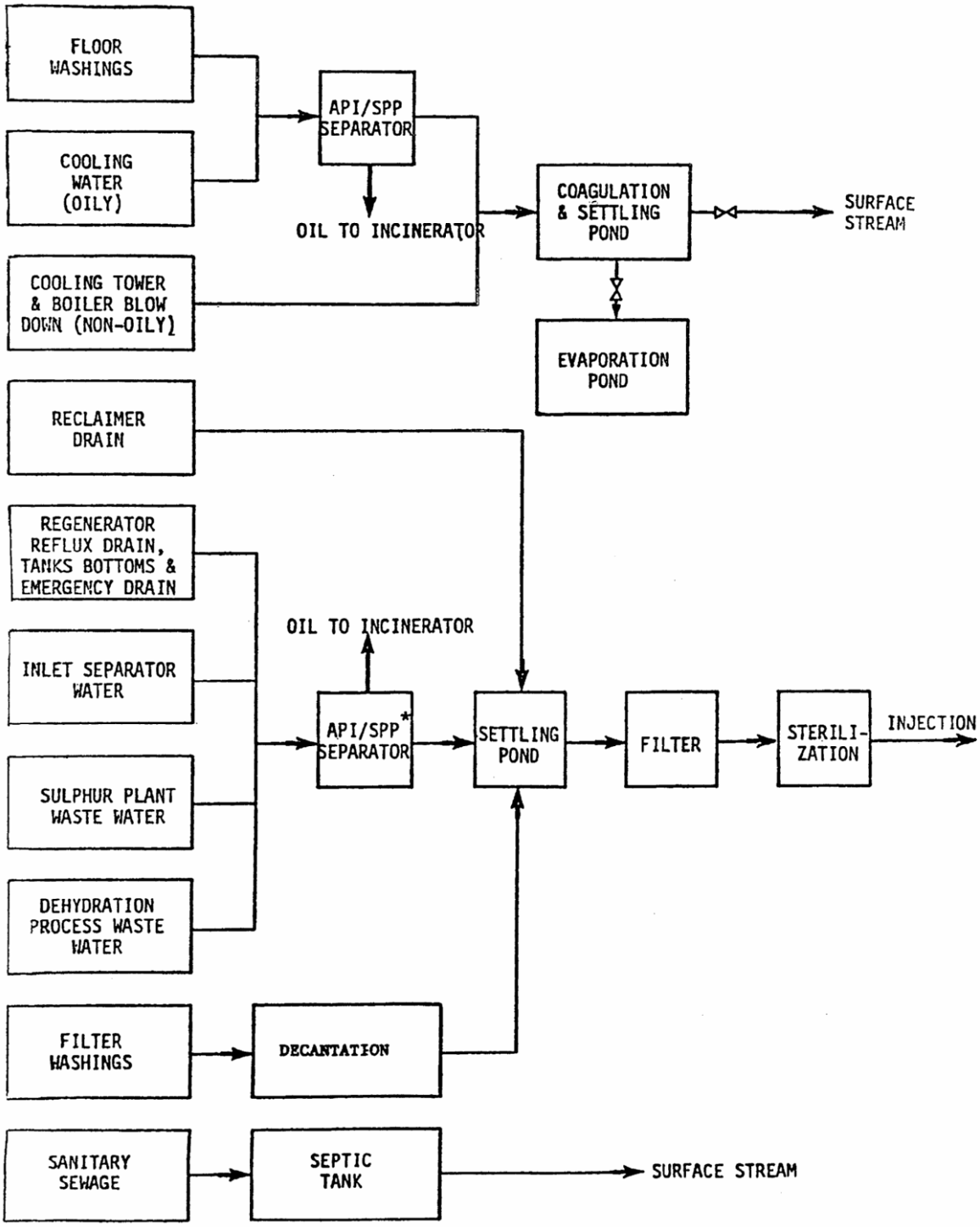


Figure 22.7: Segregation & Disposal System 1

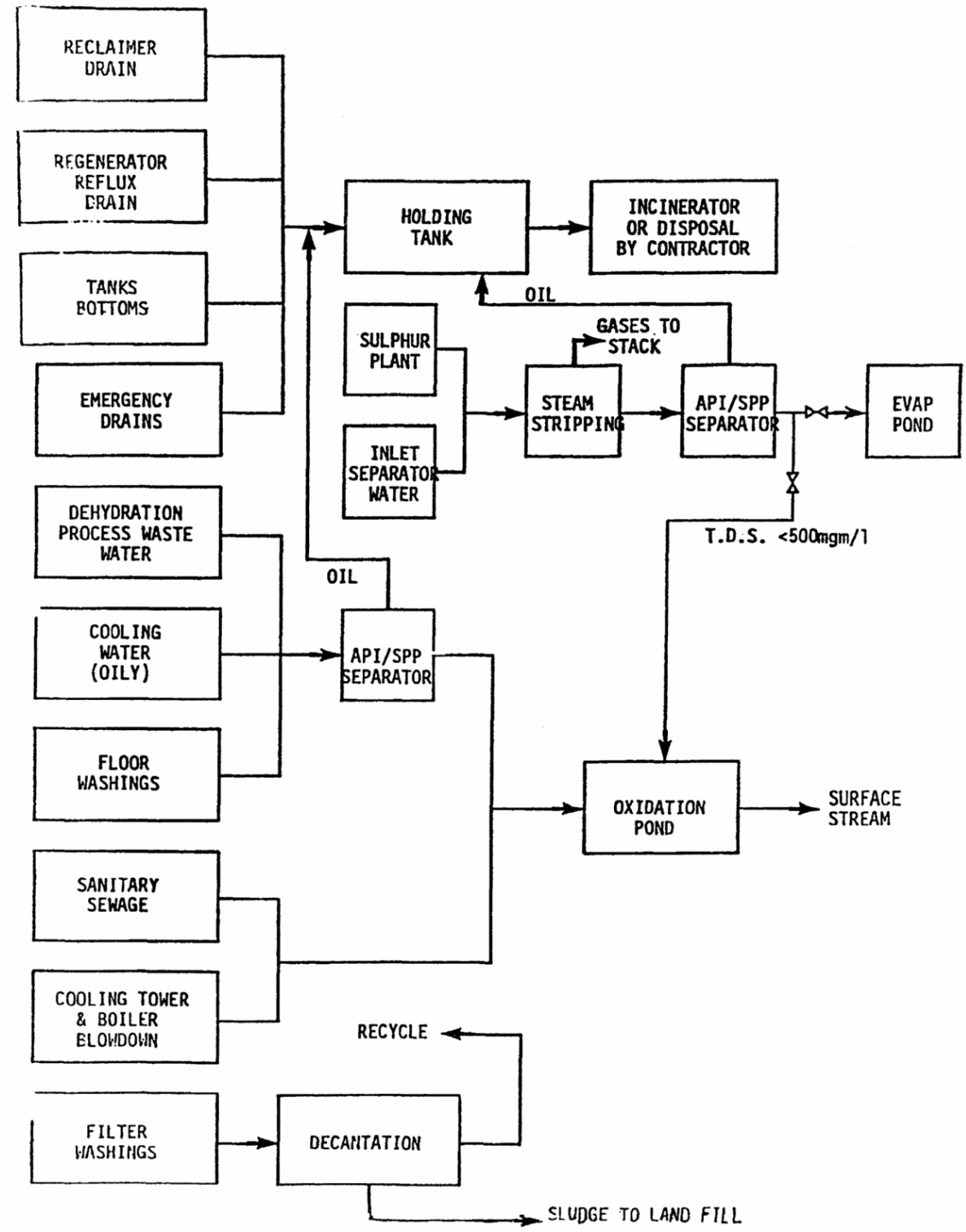


Figure 22.8: Segregation & Disposal System 2

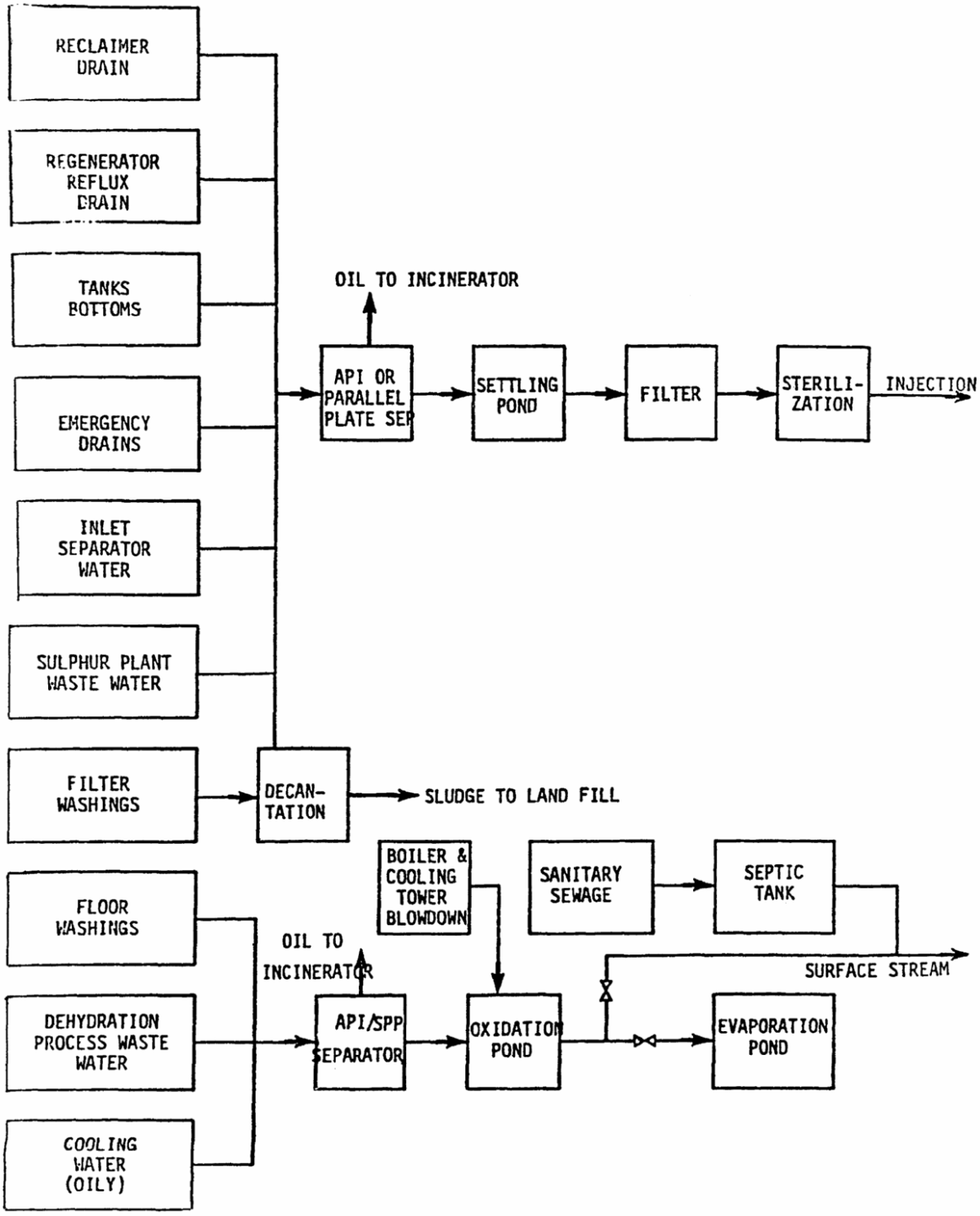


Figure 22.9: Segregation & Disposal System 3

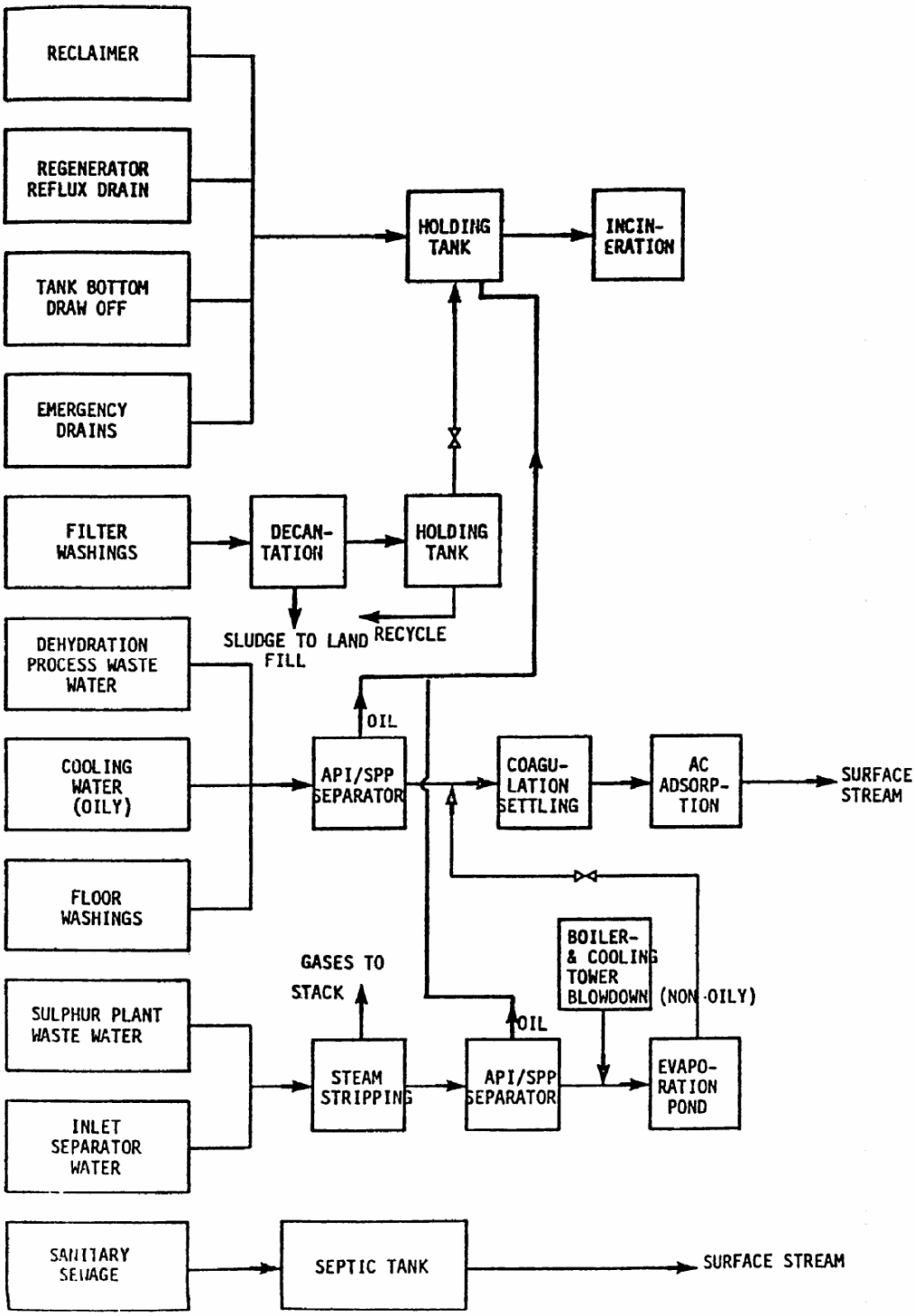


Figure 22.10: Segregation & Disposal System 4

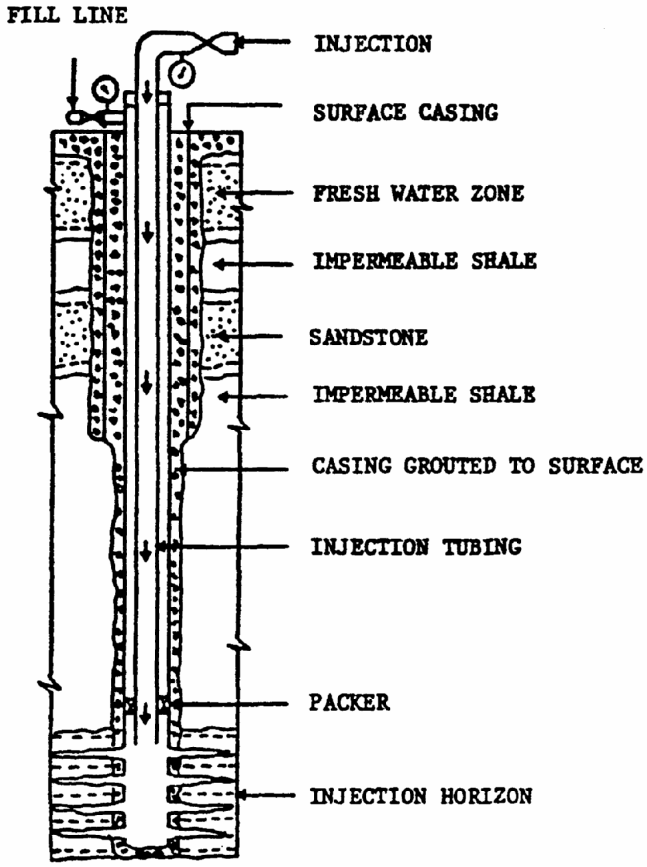


Figure 22.11: Injection Well Detail

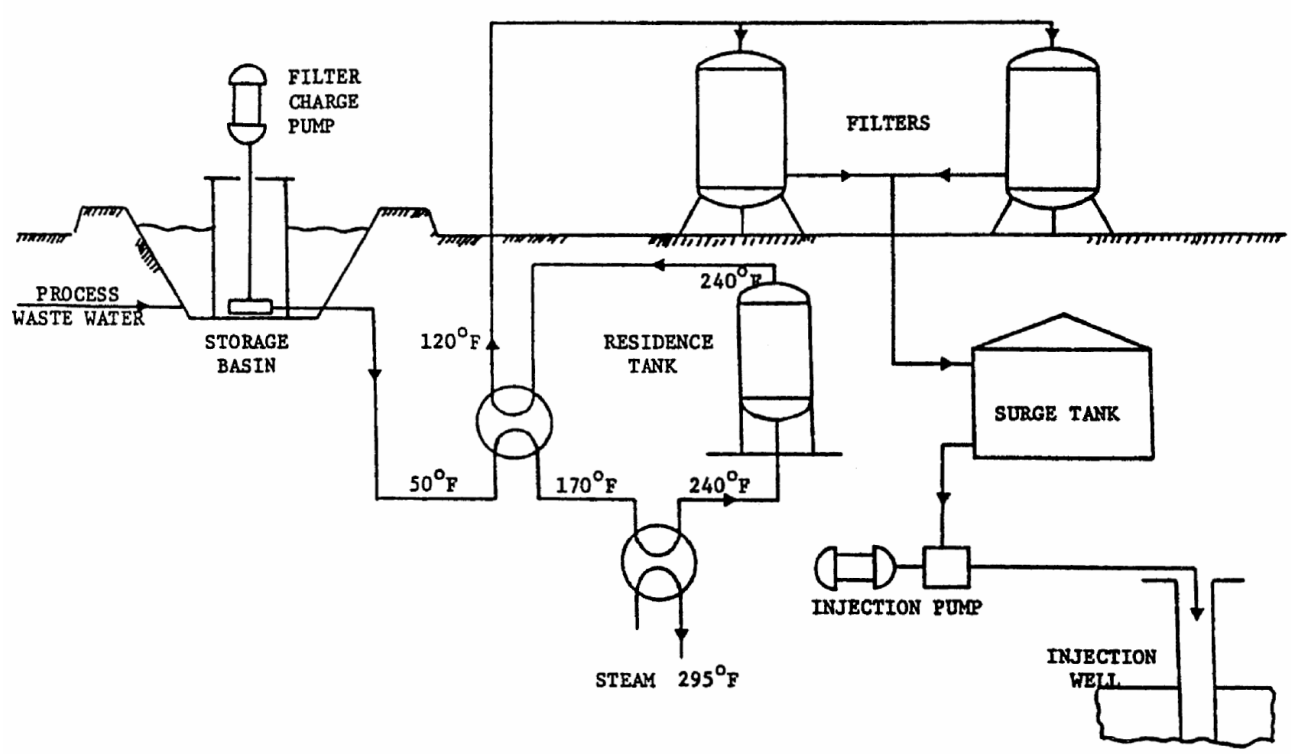


Figure 22.12: Deep Well Injection Facilities

22.4.2.2 Evaporation

As stated earlier, the quantity of aqueous wastes originating in a gas plant would depend on the size of the plant as well as on the number of unit operations employed by the plant. In certain situations the flow of effluent could be so little that its disposal by complete evaporation might be feasible. A number of gas processing plants in Alberta use shallow solar evaporation ponds to dispose of some of their aqueous wastes.

There are two major considerations in evaporation pond design: (a) the non-volatile components of the waste accumulate in the pond and need to be disposed of by alternate methods, and (b) the bottom of the pond needs to be made impermeable so that the pollutants will not seep out and contaminate the surrounding surface or underground source of fresh water.

22.4.2.3 Release to Surface Streams

The existing effluent quality forbids the release of formation-water[†] to bodies of fresh waster. Other process waste streams, however, can be released to the surrounding watershed area provided that the concentration and quantity of contaminants in the discharged waste waters do not exceed the maximum permissible levels. As indicated in Table 22.8, the following major parameters are used to monitor the effluent quality from gas processing plants:

- Chemical oxygen demand.
- Suspended solids.
- Sulphide (total).
- Ammonia nitrogen.
- Oil and grease.
- Threshold odour number.
- pH.

22.4.3 Water Quality Parameters

Various treatment methods, that can be used to remove the above mentioned pollutants, are discussed in the following paragraphs:

22.4.3.1 Chemical Oxygen Demand (C.O.D.)

This parameter indicates the presence of compounds which can be oxidized by boiling their dilute aqueous solution with chromic acid. Since most oxidizable chemical compounds can be oxidized under these conditions, the C.O.D. is expected to indicate the presence of oxygen consuming compounds better than the conventional "biological oxygen demand (B.O.D.)"[‡] test.

C.O.D. could be caused by soluble or suspended contaminants, and these could be organic or inorganic in nature. The removal of suspended matter and sulphides, which can cause high C.O.D., is discussed elsewhere. The major source of C.O.D. in gas processing plant aqueous waste are the dissolved organic compounds. Among these, the most important are sweetening chemicals and the degradation products of sweetening chemicals. These compounds are relatively resistent to biodegradation and, therefore, biological oxidation, either in aerated lagoons or

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[†] Water produced with the gas or oil and having high concentration of dissolved salts.

[‡] B.O.D. would indicate the presence of biodegradable contaminants only.

activated sludge plants, does not appear to be a suitable method for removing the C.O.D. caused by them. Also, the extreme solubility of these compounds in water makes their removal by activated carbon absorption relatively inefficient and costly. Zaidi and Tollefson suggest segregation of aqueous waste streams containing high concentrations of sweetening chemicals and degradation products (ie. filter washings, reclaimer drain and regenerator reflux drain), and disposal of these streams by incineration.

22.4.3.2 Suspended Solids (S.S.)

Suspended solids would be released with floor washings, cooling tower and boiler blowdown, and water treatment backwash. Also, the waste water stream from a dehydration plant using molecular sieves can be highly turbid.

Coagulation and settling can be used to remove the suspended solids from gas plant aqueous wastes. Lime and alum have been traditionally used as coagulants by most of the gas processing plants in Alberta to remove S.S. from the aqueous wastes. However, data reported by Zaidi and Tollefson indicate that the combination of ferric chloride and lime would be a more efficient coagulant.

22.4.3.3 Sulphides

The major source of sulphides in the gas plant effluent in the waste water stream from the sulphur plant. The concentration of sulphides in the sulphur plant waste water stream can be as high as 3 500 ppm and more. More than 99 per cent of these sulphides can be removed by steam-stripping the sulphur plant waste stream at a pH of approximately five or lower. Last traces of sulphides might be removed by oxidation with hydrogen peroxide.

22.4.3.4 Ammonia Nitrogen

This is usually accepted as an indication of sanitary pollution when encountered in raw surface waters. In the gas processing plant process effluent, the presence of alkanolamines and other organic nitrogen compounds can cause interference and result in relatively higher ammonia nitrogen levels. Segregation of sanitary effluent and the aqueous waste streams containing organic nitrogen compounds, from the rest of the effluent should help in reducing ammonia nitrogen levels to below that permissible under the present regulatory agency standards. A high level of ammonia nitrogen in cooling tower and boiler blowdown could sometimes be traced to the source of the water supply. Aeration and stripping at high pH levels can in some situations, significantly reduce ammonia nitrogen in the aqueous wastes.

22.4.3.5 Oil and Grease

The inlet separator water, the dehydration process waste water and the floor washings can contain significant amounts of oil and grease in emulsion as well as immiscible forms.

The immiscible oil can be removed from aqueous wastes by the use of either "API Separators" or "Shell Parallel Plate Separators". Both of these systems make use of the difference in the specific gravities of the two phases to separate the immiscible oil phase from the aqueous phase. Other methods, such as the absorption in polyurethane foam and froth flotation, may be required in some special situations, but generally the existing standards with regard to oil and grease content in gas plant effluent can be met with the help of properly designed API or parallel plate separators. Figure 22.13 is a drawing of a typical API separator. It is designed to provide a maximum velocity for the oily water through the horizontal concrete tank and there is a minimum residence time. The oil is removed by adjusting the skimmer into a side tank and it is recycled to the process.

The Shell parallel plate separator is claimed to be considerably more effective than the API separator and requires less space. While the API separator will remove most of the oil globules of

120 microns diameter, the parallel plate separator removes globules down to about 50 microns diameter.

The parallel plate settler contains a whole series of longitudinal parallel plates at a 450 angle, the horizontal. The oil collects on the underside of the plates and eventually rises to the surface. The introduction of the plates increases extensively the surface for the oil to coagulate on. Figure 22.14 shows the details of such a separator.

The Shell separator takes approximately 1/5 of the area of an API separator for the same separation.

If the above separation techniques are not sufficient to meet standards required for discharge into a stream, air flotation may be advantageous. This is conducted by dissolving air in the waste water at a pressure of about 50 psig and then releasing the air at atmospheric pressure in a flotation tank. The dissolved air under lower pressure comes out of solution forming tiny bubbles which rise and carry the oil to the surface. Suspended solids may be carried to the surface in the same way. The surface is continuously skimmed to recover the oil and to remove any solids which may have risen. In some instances it has been found advantageous to add about 20 ppm of alum to aid in the separation of the oil and suspended matter. This produces a floc which helps to float the oil and other suspended particulate matter.

Chemical coagulation also may be employed to treat oily and waxy paraffinic wastes. Generally this is preceded by some form of gravimetric separator and treatment with acids or baess to adjust the pH to the desired value for coagulation. The coagulants often used are alum or ferrous sulphate. The coagulant is first rapidly mixed with the waste water. In a few minutes a microscopic floc is produced which traps the oil globules and suspended solids. The flocs are allowed to slowly agglomerate with gentle mixing and finally are allowed to settle. Effluents containing no more than 20-30 ppm of oil may be decanted away from the settled sludge.

The settled sludge may be disposed of in a so-called “sanitary landfill” or it may be burned in a fluidized-bed incinerator. This latter consists of a bed of salt, mainly sodium carbonate, or sand which is maintained at about 1,300ºF initially by means of a gas burner but may require no additional fuel if there is sufficient oil present in the sludge added slowly to the bed. Combustion by the process is said to be quite complete, no air pollution problem being created.

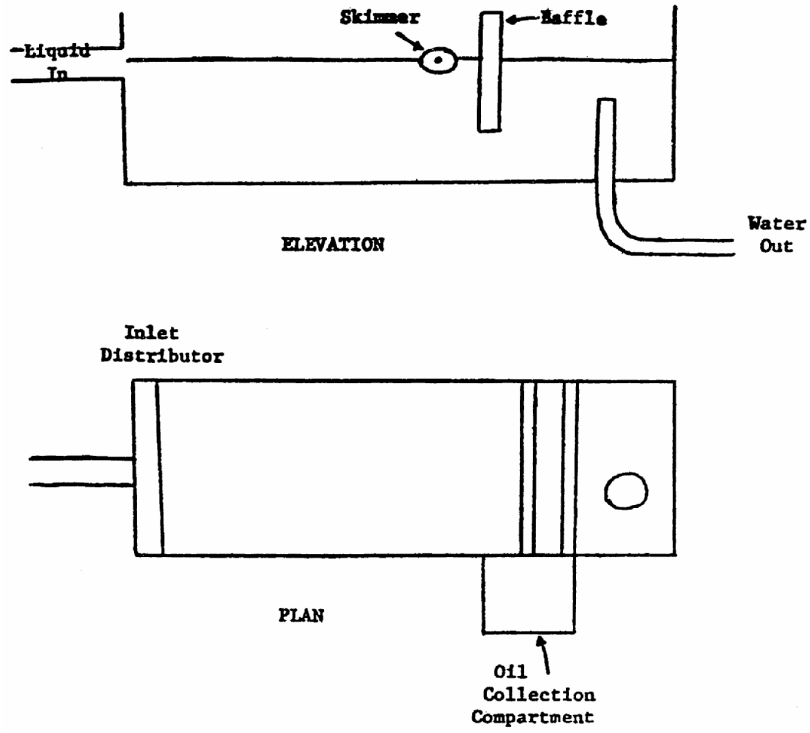

Figure 22.13: API Separator

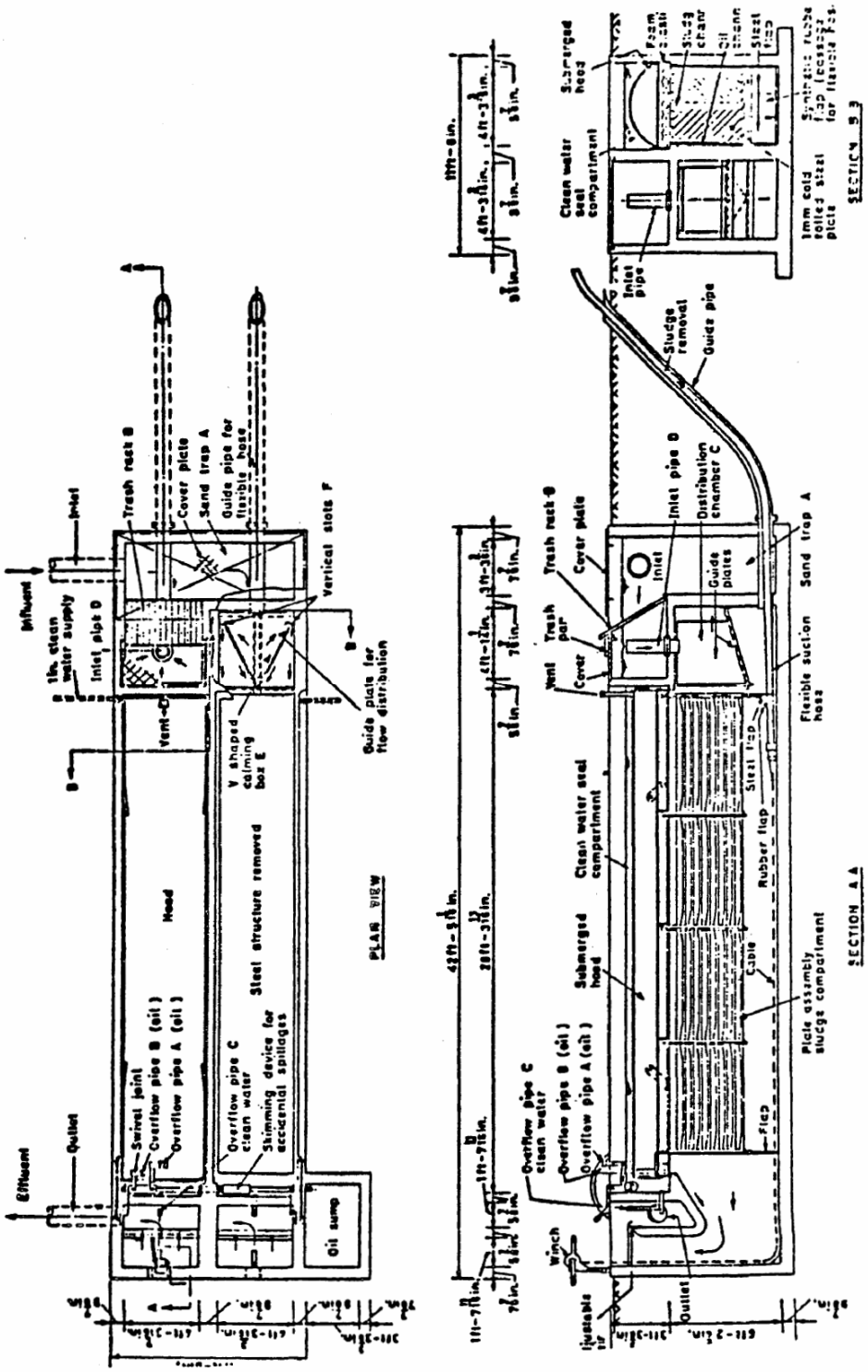

Figure 22.14: The Parallel Plate Interceptor

22.4.3.6 Threshold Odour Number (T.O.N.)

Proper removal of the aforementioned pollutants would, in most cases, reduce the T.O.N. of gas plant effluents to that permissible under the present government regulations. Special odour problems may require the use of activated carbon which could be applied in either powdered or granular form.

22.4.3.7 pH

pH might be corrected by the use of lime or sulphuric acid. Usually, the combined effluent from a gas plant would have a pH in the permissible range and it would seldom be necessary to adjust the pH.

22.5 OTHER POLLUTION CONSIDERATIONS

22.5.1.1 Disposal of Spent Amine

Amine which is used for H_2S removal may have to be discarded for various reasons. It should be incinerated in either a furnace equipped with a stack or in a bed of hot fluidized solids such as was described earlier. If these are not available it may be injected into a deep disposal well. Normally considerable effort should be expended in recovery of amine by other processes since wasting them represents considerable economic loss.

22.5.1.2 Disposal of Spent Caustic

Caustic soda may be used to remove trace quantities of H_2S and mercaptans from gas streams. Generally it is regenerated by air oxidation but stream stripping may also be used. In any case, some spent caustic must be disposed of. This may be burned in an incinerator or a fluid bed reactor to produce Na_2CO_3 or it may be sold sometimes. Some SO_2 and other S-containing gases may be given off and these should be handled via a flare stack. If it cannot be sold, it is preferable to dispose of the spent caustic by injecting it into a disposal well.

22.5.1.3 Thermal Pollution

This is not normally a problem in Alberta. Rapid changes in the temperature of the water of a river can cause shock to the aquatic life in the river. For this reason, the water temperature should not be increased by more than 3ºC above the ambient water temperature as the result of discharge of hot waste water.

22.5.1.4 Liquid Disposal from Gas Well

When water is produced from a well and must be disposed of without the aid of permanently installed equipment, an earthen pit may be used for storage according to 'Board' regulations. The pit is to be constructed of clay or other suitable impermeable material. Its surface area is to be limited to 600 square feet. It is to have walls rising two feet above ground level and should be dug to the depth that is required to contain the waste water. Not more than 100 barrels per month normally can be stored or disposed of in this manner. In some instances, water stored in this manner slowly evaporates or seeps slowly into the ground. Where this does not happen, injection of the waste water into an underground formation is a desirable way to dispose of the material provided approval of the Board is obtained for so doing.

22.5.1.5 Soil Pollution

Loading of sulphur on windy days can cause soil pH near the sulphur loading area to decrease significantly, a decrease from a value of 6.0 to between 2.5 and 3.5 having been observed in one operation.

Lime addition can be used to neutralize this acidic condition but the salts produced form liming tend to accumulate in the soil and alter its permeability. Regulatory agencies are now requiring sulphur handling methods that minimize sulphur dust. This has led to the development of processes for making prilled, pelletized and slate sulphur for bulk handling of the solid material. This is discussed in the Sulphur Plant Section.

Soil pollution may also result for salt water spills, and especially form leakage in water disposal lines or in the vicinity of salt water injection wells and trunk unloading areas. Care must be exercised in the design and operations of salt water handling facilities to minimize such contamination.

22.6 Noise Pollution

Currently, the only legislation applicable to gas processing operations is enacted under the Public Health Act and it is enforced by the Worker's Compensation Board. This provides that ear protection must be supplied for all employees working in areas where the noise level exceeds 85 dBA.

The ERCB has issued interim directive No. ID-80-2 as a guideline on noise. The directive states that 15 meters from a residence on the side nearest the noise source, the noise level generally must not exceed 65 dBA during the day and 50 dBA at night between 11:00 p.m. and 7:00 a.m. A revised directive is to be issued shortly.

As an introduction to the topic of noise measurement, consider the quesiton - What is noise? It is unwanted sound. With this definition, it is very difficult to precisely classify sound because people respond differently to sound. However, it is possible to have people with good hearing state when a sound is barely heard, and when it is painful to the ear. With a sensitive microphone to measure the pressure level of the sound waves electronically, it will be found that the painful noise pressure level is about 1,000,000 times stronger than the smallest level sensed by the human ear, which for convenience, has been assigned a sound pressure of 1. If the logarithmic scale is to the base 10, then the barely audible level becomes 0 and the painful level becomes 6.

Converting sound pressure level to the energy units, all of the sound pressure levels must be squared. The range of sound would then be 1 to 1 x 10^{12}; or logarithmically, 0 to 12. After Alexander Graham Bell, this unit of sound pressure level scale was called "bels" and for accuracy the unit was divided into 10 smaller units, called logically, the decibel, so that the scale of sound level would now extend from 0 to 120 decibels, or the dB. Thus, for every 10 dB, the sound pressure level doubles.

Since the logarithmic scale is used in dealing with noise problems, it can be somewhat confusing when noise sources are added together or when correction of noise problems is being worked on. Consider two noise sources which are exactly similar, operating together. The sound energy generated by the two when added together will be double that of each source operating along.

[¥]However, the sound pressure level in dB from the combined sources will only be 3dB higher than the level produced by either operating alone. Two noise sources which differ in sound pressure level by 10dB or more, when added together, will result in a level not significantly different than that

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[¥] Sound level x + Sound level x + Sound level 2x = 10 log (2x)
= 10 log 2 + 10 log x
= 3 + 10 log x

for the loudest operating alone. Thus, the noisiest source must be reduced to achieve a significant reduction in noise.

There have been many scales developed to physically relate sound pressure levels to a basis which reflects human perception and reaction. The most commonly used scale for ground transportation, industrial and community noise, is the A-weighted sound level, dBA.

It is well-known that the sound pressure levels vary over the audible frequency range. Low frequencies cannot be heard as readily as most of the higher frequencies. Thus, through a system of sensitive microphones, it is possible to. adjust the sound according to the general loudness sensitivity in each frequency band and add the contribution of each to give a single reading in dBA. The dBA scale in effect, is a weighting of the sound pressure level at various frequencies which is representative of the hearing capability of the human ear.

22.7 Final Draft Noise Guidelines ERCB Noise Committee

Introduction

The procedures for calculating noise levels for both design and compliance situations are outlined below. The Design Noise Level and Compliance Noise Level are the composite of all noises, including background noise, traffic noise, and industrial noises. For both design and compliance situations, the Basic Noise Level is utilized.

Basic Noise Level (see Table 1)

The Basic Noise Level is representative of background noise levels due to local activities characteristic of the area, plus an allowance for industrial presence.

Design Noise Level

For facility design purposes, the Design Noise Level represents the maximum allowable noise level due to the facility. It shall apply 15 m from the nearest or most impacted dwelling unit. The Design Noise Level is derived from a base value intended to be representative for the type of area plus adjustments intended to more accurately reflect specific aspects of the facility and the environment, and shall be calculated as follows:

$$Design\ Noise\ Level = Basic\ Noise\ Level + Class\ A\ Adjustment + Class\ B\ Adjustment$$

The adjustments to the Basic Noise Levels shall not create Design Noise Levels less than 40 dBA Leq or more than 70 dBA Leq during the appropriate time periods.

Compliance Noise Level

For compliance purposes a facility shall meet the Compliance Noise Level as determined by noise monitoring, which shall be performed when a complaint is received or at the request of the Board. The Compliance Noise Level shall apply 15 m from the complainant's dwelling unit. It shall be calculated from the Measured Ambient Noise Level obtained from a compliance noise monitoring survey minus adjustments intended to more accurately reflect specific aspects of the facility and the environment. The Compliance Noise Level shall be calculated as follows:

Compliance is then achieved when the Compliance Noise Level, rounded to the nearest whole number, does not exceed the Basic Noise Level (Table 1).

$$Compliance\ Noise\ Level = Measured\ Ambient\ Noise\ Level - Class\ A\ Adjustment - Class\ B\ Adjustment$$

The Measured Ambient Noise Level shall be determined by conducting a noise monitoring survey. The survey shall be conducted continuously for a minimum 6 hour period to a maximum 24 hour period. The survey shall encompass a representative portion of the times of day or night on typical days when the noise causing the complaint occurs, and shall be based upon the nature of the noisy operation and environmental conditions. If the required survey period straddles the daytime/night time periods, then a minimum 6 hour monitoring period with at least 3 hours within both the daytime and night time periods shall be conducted.

When it appears that the facility's noise contribution is below the ambient noise level, the facility's contribution can be determined by using an isolation analysis technique acceptable to the ERCB. In such cases, the isolated facility's contribution could be used in place of the Measured Ambient Noise Level.

The Compliance Noise Levels do not apply for emergency situations. An emergency is defined as an unplanned event requiring immediate action to prevent loss of life or property.

Measurement Instrumentation

The sound measurement instrumentation to conduct the noise monitoring survey is required to measure the A-weighted (dBA) continuous energy equivalent sound level (Leq) of steady, intermittent, and fluctuating sounds. The instrumentation must be able to accumulate the data and calculate the Leq's over the time periods required herein. The instrumentation used must meet the minimum technical specifications in International Electrotechnical Commission (IEC) publication #804 for Type II sound level meters.

The sound measurement instrumentation necessary to conduct the 1/3 octave band sound pressure level measurements to characterize the presence of tonal components must meet the minimum technical specification in IEC publication #225-1966 or American National Standards Institute (ANSI) publication #S1.11-1966 for Class II filter sets, used in conjunction with conventional sound level meters which meet the minimum technical specifications in IEC publication #651-1979 or ANSI publication #SI.4-1983 for Type II sound level meters.

The sound measurement instrumentation necessary to conduct the impulse sound level measurements to characterize the presence of impulse/impact components must meet the minimum technical specifications in IEC publication #651-1979 or ANSI publication #S1.4-1983 for Type II sound level meters.

Better quality instrumentation may meet many or all of the specifications mentioned herein.

Measurement Techniques

The sound measurement techniques employed must be carefully chosen and controlled to obtain valid and consistent results. This includes the effects of environmental factors, the presence of the operator and the local setting, etc. General guidelines for sound measurement techniques can be obtained from excerpts of the following publications: "A Method for Conducting and Reporting Noise Surveys at Industrial Plants", Alberta Environment; "Model Municipal Noise Control By-Law", Ontario Ministry of the Environment; "Methods for the Measurement of Sound Pressure Levels", ANSI publication #S1.13-1971; and "Assessment of Noise with Respect to Community Response", International Organization for Standardization (ISO) publication #1996; also sound measurement manuals and instrument instruction manuals as available from instrumentation manufacturers.

For the convenience of the reader, the following are the more important considerations which require attention:

- Calibration - acoustic calibration must be performed before and after each survey.
- Microphone position and orientation - the microphone must be a minimum of 1.2 m above the ground. Use of a tripod is required. The microphone must be a minimum of 3 m away from significant sound reflecting surfaces, and oriented as per instrumentation manufacturer's instructions. Use of approved windscreen required.
- Wind noise - hourly average wind speeds greater than 15 km/hr invalidate data obtained during the affected period. Short term wind gusts up to 20 km/hr are acceptable.
- Rain - invalid for monitoring.
- Abnormal noise events - potentially invalid for monitoring.

In cases where a discrepancy occurs between measurement techniques mentioned herein and those presented in the cited references, these guidelines shall prevail. The Board reserves the right to pass judgement regarding the suitability of any sound measurement techniques employed.

Table 22.11: Basic Noise Levels

The Basic Noise Levels are representative background noise levels due to local activities characteristic of the area, plus an allowance for industrial presence.

	Dwelling Unit Density Per Section of Land					
	1 - 32 Dwellings		**33 - 640 Dwellings**		**640 Dwellings**	
Proximity To Transportation	*Daytime Level*	*Night Time Level*	*Daytime Level*	*Night Time Level*	*Daytime Level*	*Night Time Level*
Category 1	50	40	55	45	60	50
Category 2	55	45	60	50	65	55
Category 3	60	50	65	55	70	60
Daytime Level	A-weighted, continuous energy equivalent sound level from 0700 to 2200 hr (dBA Leq).					
Night Time Level	A-weighted, continuous energy equivalent sound level from 2200 to 0700 hr (dBA Leq).					
Category 1	Dwelling units more than 500 m from heavily travelled roads and/or rail lines, and not subject to frequent aircraft flyovers.					
Category 2	Dwelling units more than 30 m but less than 500 m from heavily travelled roads and/or rail lines, and not subject to frequent aircraft flyovers.					
Category 3	Dwelling units less than 30 m from heavily travelled roads and/or rail lines, and/or subject to frequent aircraft flyovers.					
Dwelling Unit	Any permanently or seasonally inhabited residence.					
Density Per Section	Refers to the affected inhabited section. For sections with various land use or with mixed densities, the density chosen shall be prorated for the land parcel area under consideration.					
Heavily Travelled Roads	Includes Primary and Secondary roads as designated by Alberta Transportation, or major roadways as designated by the appropriate authorities.					
Rail Lines	Includes Main and Branch lines as designated by the appropriate authorities.					

Table 22.12: Noise Level Adjustments

These adjustment values are used to establish more appropriate Design Noise Levels and Compliance Noise Levels, and are based upon the nature of the activity and peoples response to noise.

Class	**Adjustment**[1]	**Value (dBA Leq)**
A1	Seasonal Adjustment (Nov 1 - Mar 31)	5
A2	Heavy Truck Adjustment[2]	0 to 5
A3	Absence of both Tonal and Impulse/Impact Components Adjustment[3]	5
A4	Ambient Monitoring Adjustment[4]	- 5 to 5
Class A Adjustment = Sum of A1, A2, A3, and A4 (if applicable), but not to exceed a maximum of 10 dBA Leq.		
Temporary Activity Adjustment for total activity complete within:		
B1	1 day	15
B2	1 week[5]	10
B3	1 month[5]	5
B4	greater than 1 month[5] or no temporary activity	0

Class B Adjustment = One only of B1, B2, B3, or B4.
[1] Applicable only upon submittal of technical verification.
[2] Applicable only within 1 km radius of the facility.
[3] Useable only when audible characteristics of permanent facility are absent of both commonly annoying Tonal and Impulse/Impact Components. Absence of tonal component shall be when the measured 1/3 octave band, slow response, Sound Pressure Levels of any one or two adjacent band(s) between 20 Hz and 16,000 Hz that form the high points of a spectrum is always 10 dB or less above the Sound Pressure Levels in bands adjacent to these bands. Absence of impulse/impact component shall be when the difference between the A-weighted impulse response sound level measurement and the A-weighted slow response sound level measurement is 10 dBA or less.
[4] Useable only when Basic Noise Levels (Table 1) may not be representative of actual ambient noise environment. Applicable only when ambient noise monitoring survey has been conducted prior to application approval. Ambient noise monitoring survey shall consist of a 24 hour continuous noise monitoring survey, with measured ambient noise levels presented for the daytime and night time periods, conducted 15 m from the nearest or most impacted dwelling unit. If Adjustment A4 is claimed during the application process, it must also be used when determining compliance.
[5] Cumulative duration (in days) for periods of noisier operation(s) during temporary activity. Adjustment values applicable for day(s) with lowest anticipated activity.

Table 22.13: Heavy Truck Adjustment - A2

This adjustment reflects the influence of vehicular traffic's noise contribution. As it is recognized that truck noise is more tolerable than facility noise, an adjustment of noise levels is permitted up to a maximum 5 dBA Leq, based upon the amount of heavy traffic that exists.

Distance To Roadway	Daytime (0700-2200 hrs)						Night Time (2200-0700 hrs)					
	Basic Noise Level[1]	Adjustment[2]					Basic Noise Level[1]	Adjustment[2]				
		1	2	3	4	5		1	2	3	4	5
Less Than 30 m		Heavy Truck[3] Passbys[4] in Daytime Period						Heavy Truck[3] Passbys[4] in Night Time Period				
							40		1	2	3	4
							45	6	8	10	12	15
	50	24	30	38	48	60	50	15	20	25	30	40
	55	70	85	110	135	170						
	60	160	200	250	320	400						
More Than 30 M, Less Than 100 M		Heavy Truck[3] Passbys[4] in Daytime Period						Heavy Truck[3] Passbys[4] in Night Time Period				
							40	10	13	16	20	25
							45	25	30	40	50	60
	50	120	150	185	235	300						
	55	320	400	510	640	800						

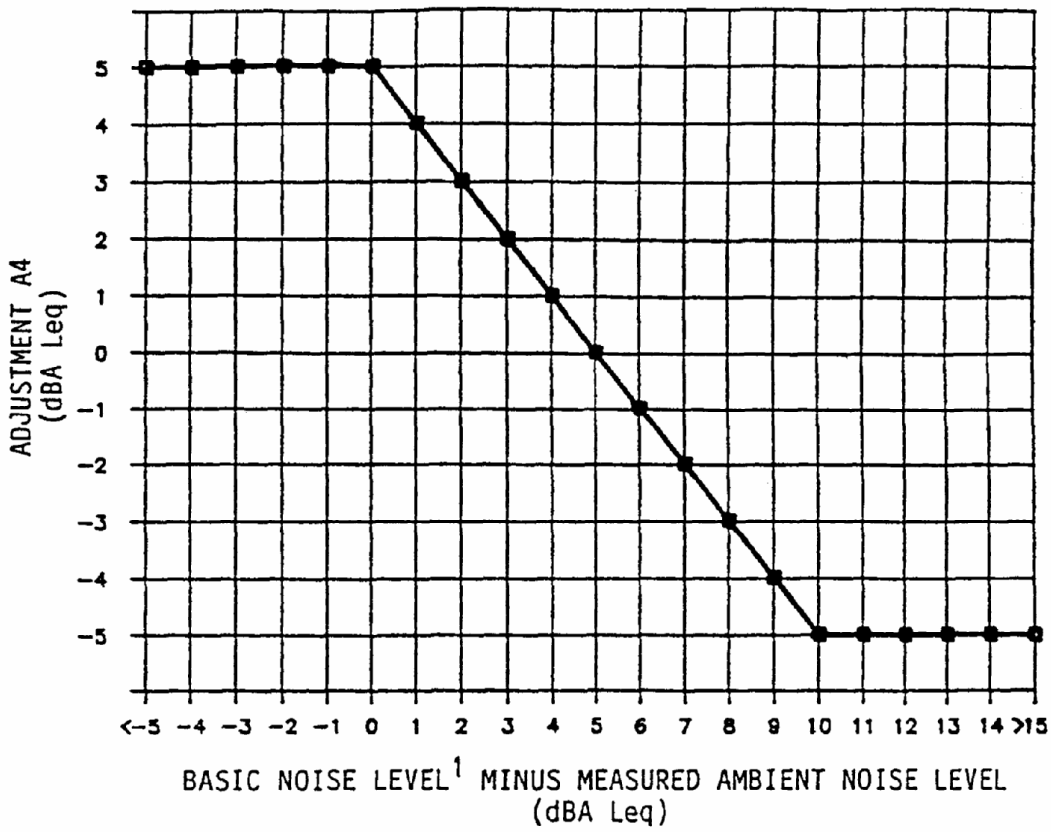
[1] From Table 22.11

[2] Adjustment A2 to Basic Noise Level (dBA Leq) based upon number of passbys.

[3] Any energy-facility related truck greater than 12,000 kg GVW or having three or more axles.

[4] For number of heavy truck passbys falling between values, use next highest passby value. Note that a round trip may include two passbys. When truck counts are taken for compliance purposes, if the monitoring period is less than the full 15 hour daytime or 9 hour night time measurement period, the number of truck passbys shall be prorated as follows:

$$Prorated\ \ passbys = Counted\ \ passbys \times \frac{Actual\ measurement\ \ period\ (hrs)}{Full\ 15\ hr\ or\ 9\ hr\ measurement\ \ period\ (hrs)}$$



[1] from Table 22.13

Figure 22.15: Ambient Monitoring Adjustment - A4

This adjustment allows for an incremental change of the Basic Noise Level to reflect characteristics of the actual ambient noise environment.

22.8 Problems

Example Problem 22.1

A flare 100 ft. high is discharging 150 MMSCFD (assume 700 as standard conditions) of gas a day and it suddenly goes out. A man is welding on the ground relatively close by. Is he in particular danger from an explosion? The velocity of the gas leaving the stack is 100 ft./sec. The wind velocity at the time is 10 mph.

Example Problem 22.2

A sulphur plant has a fire at the top of its 300 ft. stack. It is burning approximately 200 tons/day in this fire as the sulphur plant is not working. The time is winter with the air temperature at 0ºF. The temperature of the fire is 2000ºF. Assume the wind velocity in 20 miles per hour. The calculated thermal rise for 1000ºF exit temperature and 60ºF surrounding temperature is 100 ft.

Calculate the ground level concentration relatively near the stack.

Example Problem 22.3

The first converter of a 3 converter sulphur plant is taken out of service because of a grating failure ? to change the catalyst. It will be out of line for one week. However, because of gas requirements

it is necessary to keep the plant operating. You are to determine the ground level concentration 1km (3255 ft.) from the stack when operating with the first converter down.

The plant also has a sulfreen unit that will recover the sulphur in the tail gas so that the total plant will meet the ERCB emission regulations with the Clause unit working at 94% recovery.

Gas flow 100 MMSCFD acid gas

Gas Composition: 70% H_2S

28% CO_2

2% H_2

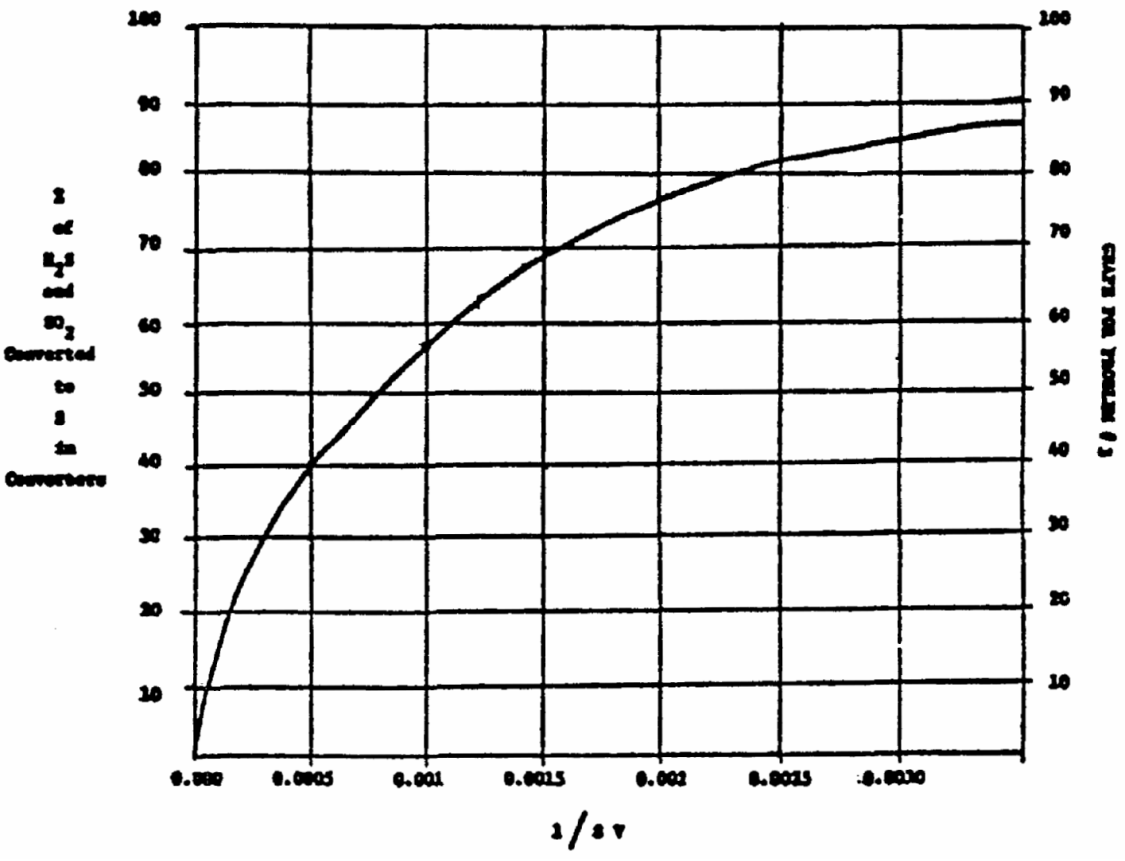
Converters contain 12,000 cubic feet of catalyst.

Assume 20% conversion in the reaction furnace.

The graph shown below gives the approximate conversion expected in the converters.

Data

- The height of the incinerator stack is 450 ft.
- The surrounding terrain is flat.
- The height of the trees locally is 50 feet.
- The buoyancy flux F is 3200 ft. sec^{-3}.
- The wind velocity at the time is assumed to 20 mph (29.34 ft./sec.).



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23.0 Cryogenics

- A general review of the special features of handling gas and its products at low temperatures.
- This revision has been prepared with the help of K. C. Milne.

23.1 What is Cryogenics?

Cryogenics is generally considered the study of processes and the study of the properties of materials at low temperatures. Cryogenic processing is probably considered to start at -100ºF (-73ºC) while ordinary low temperature processing is considered to be around -50ºF. Processes in the area of -100ºF to -150ºF (-73ºC to -101ºC) are the start of the cryogenic range. We are interested in cryogenics particularly because of the manufacture of liquefied natural gas (LNG) although the principles are applied in the manufacture of several industrial gases such as oxygen, nitrogen, helium, etc. LNG is made at places with a surplus of gas and shipped in specially designed tankers to consumers around the world. LNG is also used for peak shaving, ie. it is produced and stored at times of surplus gas, such as in the summer, and revaporized in the winter. Methane, which is the principal component of LNG, is a liquid at -259ºF (-161ºC) at atmospheric pressure.

23.2 General

In the field of natural gas processing, cryogenics is used in NGL recovery processes for the production of ethane and propane plus that operate in the range of -100ºF to -150ºF such as is used in the Strachan, Waterton, Bonnie Glen, Empress, Edmonton, Cochrane, Jumping Pound, and Garrington plants in Alberta. There is probably over 6 BSCFD of gas processed in Alberta using the cryogenic turboexpander process. This process has been discussed in some detail in Chapter 19. There are other cryogenic plants that are special situations such as the rejection of N_2 such as is done at Willisden Green.

Cryogenics that we will consider here is production of LNG such as that proposed by TransCanada Pipelines and Petro-Canada for shipment from the Arctic, and by Dome for shipment to Japan. These processes use operating temperatures down to -260F (-162C). In Canada, at the present time, there are three LNG peak shaving plants having a total capacity of 20 MMCFD (0.57×10^6 m^3/d). These are located at Vancouver, at Hagar in Ontario, and on Montreal Island. In the base load plant size (plants liquefying over 50 MMSCFD - 1.4×10^6 m^3/d) there are plants around the world with a design capacity of about 13,000 MMSCFD (366×10^6 m^3/d). The only North American base load plant is that of Phillips located in Kenai, Alaska. Table 23.1 lists the base load plants in operation and their capacity. Reference 16 gives an update (till the end of 1986) on the industry.

23.3 Physical Data

The physical data required for cryogenic design are:

1. Thermodynamic properties.
2. Physical properties.
3. Heat transfer coefficients.

4. Fluid flow pressure drops at the low temperatures expected, ie. at -100ºF to -275ºF (-73ºC to ~170ºC)

The thermodynamic properties of importance are the liquid and vapour equilibrium ratios, enthalpy, and entropy. Entropy of gas at low temperatures and high pressures is important primarily in the expander process.

The physical properties of importance are liquid and vapour density, viscosity, and thermal conductivity. These properties are required in order to properly size the heat exchangers. Chapters 2, 3 and 4 should be read and references 1, 2, 3, 4 and 5 reviewed.

A method for calculating the tube side and shell side heat transfer coefficients is needed, particularly in LNG plants. The cryogenic heat exchanger in an LNG plant on tube side has both a single phase region and a two phase condensing region, while on the shell side it has a single phase region and a two phase vaporizing region. References 6 and 7 discuss this subject in some detail.

Because of the importance of pressure drop in the cryogenic heat exchanger, an accurate method for calculation of pressure drops for single and two phase flow on the shell and tube sides is required.

Both the refrigeration system and the cryogenic heat exchangers are designed by the process licensor in an LNG process. Thus, it is necessary for the licensor to have this data in-house. They have obtained this by extensive research and in plant tests and consider the information proprietary.

The expander processes are handled more like conventional natural gas processes where the thermodynamic properties are known to the process designer. But in the LNG process the physical, heat transfer, and pressure drop properties are developed by exchanger manufacturers and/or the LNG process licensor.

References 6 and 7 review this subject further.

Table 23.1: International Base-load LNG Liquefaction Facilities

Company & Plant Site	Liquefaction Plant Capacity 10^6CF/d	Storage 10^6CF/d	Capacity 1000 bbl	Type of Storage Container	Storage Contractor	Liquifaction Plant Contractor	Type of Cycle	Year of Operation
CAMEL (GLA-Z) Arsew, Algeria	200	1840 total	210 (3 x 70)	Above ground, (2 - aluminum, 1 - 9% nickel)	CBI llc.	TECHNIP	Cascade	1963
			310	Frozen Hole	Pritchard	Pritchard		
SONATRACH (GL1-Z) Arsew, Algeria	1100	7000	1890 (3 x 630)	Above ground, 9% nickel	Toyo Kanatsu K.K. (PDM Lic)	Bechtel Corp. Chemico	APCI-MCR	1978
SONATRACH (GL2-Z) Arsew, Algeria	1000	6000	1800 (3 x 600)	Above ground, 9% nickel	Toyo Kanatsu K.K. (PDM Lic)	Pullman Kellogg	APCI-MCR	1982
SONATRACH (GL1-K) Skidka, Algeria	430	2500	700 (2 x 350)	Above ground, 9% nickel	CMP/CBI	TECHNIP	TEALARC	1972
SONATRACH (GL2-K) Skidka, Algeria	170	1250	350	Above ground, 9% nickel	CBI	Pullman-Kellogg	PRICO	1981
SONATRACH (GL3-K) Skidka, Algeria	350	3500	880 (2 x 440)	Above ground, 9% nickel	CBI/Whessoe	Pullman-Kellogg	PRICO	1982
SIRTE OIL CO. Marta el Brega, Libya	385	2100	630 (2 x 315)	Above ground, 9% nickel	CBI	Esso R&E Bechtel SNAM Progetti	APCI-MCR	1970
BRUNEI LNG Lumut, Brunei, Borneo	750	4050	1131 (3 x 377)	Above ground, 9% nickel	Toyo Kanatsu K.K. (PDM Lic)	Japan Gasoline CO--UOP	APCI-MCR	1973
ABU DHABI GAS LIQUEFACTION CO. LTD., Das Island, Abu Dhabi	350	6600	1920 (2 x 960)	Above ground, 9% nickel	IHI	Bechtel-Chiyoda	APCI-MCR	1977
ABU DHABI GAS LIQUEFACTION CO. LTD., Das Island Abu Dhabi	--	5033	1510 (3 x 503)	Composite	CBI	--	--	1985; replacement for existing storage
PHILLIPS-MARATHON Kanai, Alaska	185	2300	675 (3 x 225)	Above ground, aluminum	CBI	Phillips-Bechtel	Cascade	1969
PERTAMINA/MOBIL Biang Lancang, W. Sumatra Indonesia	1200	8350	3200 (4 x 800)	Above ground, aluminum	Toyo Kanatsu K.K. (PDM Lic)	Bechtel, Inc.	APCI-MCR	1978
PERTAMINA/MOBIL Biang Lancang, W. Sumatra Indonesia (plant expansion)	550	--	--	--	--	Chyoda	APCI-MCR	1984
PERTAMINA/HUFFCO Badak, E. Kalimantan, Ind.	530	8330	2400 (4 x 600)	Above ground, 9% nickel	Hitachi Zosen- CBI	Bechtel, Inc.	APCI-MCR	1977
PERTAMINA/HUFFCO Badak, E. Kalimantan, Indonesia (plant expansion)	530	2777	800	Above ground, 9% nickel	CBI	Bechtel, Inc.	APCI-MCR	1983
MALAYSIAN LNG SDN. BHD. Bintulu, Sarawak, Malaysia	850	5690	1636 (4 x 409)	Composite	IHI	Pullman-Kellogg-Japan Gasoline Co. Ltd.	APCI-MCR	1983
WOODSIDE OFFSHORE PETROLEUM Northwest Shelf Western, Australia	370	--	1636	Above ground, 9% nickel. Earthen Berm around same			APCI-MCR	1989

23.4 Overall LNG Production

Although the gas that is normally fed to an LNG liquefaction plant meets the normal specifications for pipeline or domestic gas (see Chapter 1 for specification for this material), it is not nearly pure enough to be fed directly to the liquefaction section. Water, heavy hydrocarbon and carbon dioxide will all freeze and form solids in the low temperature section; thus they must be removed. Also, H_2S should be removed as much as possible. Another impurity that has caused trouble is mercury. The presence of this element in the natural gas should be checked for. Table 23.2 shows the allowable feed impurities in LNG.

The overall process scheme is shown in Figure 23.1. Generally, just CO_2 removal and dehydration to a dewpoint of -260ºF (-162ºC) is required.

The CO_2 and H_2S are removed by a conventional sour gas treating process, as described in some detail in Chapter 16. Generally, an amine process (mostly MEA) has been used although sulphinol and selexol have been considered. The CO_2 and H_2S are normally removed below 25 ppm.

Table 23.2: Allowable Feed Impurities Considering Solubility in LNG

Compound	(Molar)	Note
CO_2	40 ppm	(5)
H_2S	735 ppm	
Methyl Mercaptan	47 ppm	
Ethyl Mercaptan	134 ppm	
COS	3.2%	(2)
Iso-Butane	62.6%	(2)
Normal Butane	15.3%	(2)
Iso-Pentane	2.3%	
Normal Pentane	0.89%	
Hexane	217 ppm	
Heptane	70 ppm	
Octane	0.50 ppm	
Nonane	0.10 ppm	
Decane	0.51×10^{-5} ppm	
Cyclohexane	115 ppm	
Methylcyclopentane	0.575%	
Methylcyclohexane	0.335%	
Benzene	1.53 ppm	
Toluene	24.9 ppm	
O-Xylene	0.22 ppm	
M-Xylene	1.54 ppm	
P-Xylene	0.012 ppm	
H_2O	0.10×10^{-4} ppm	(3)
Mercury	----	(4)
(1) Based on LNG solubility in net storage tank LNG and corrected to feed flow, with a 1.2 factor for data uncertainty and		

a 1.1 factor for boil-off of 10%.
(2) These high concentrations would change the solvent character if present to the extent shown and would require re-calculation of other components' solubilities. This was not considered to be a likely situation, so all solubilities listed were calculated with the composition of the net LNG (stream 18) as the solvent.
(3) By experience, gas which has been dried to 0.5 ppm will not encounter solubility problems with water.
(4) Not based upon solubility. Mercury should not be allowed in the feed in any concentration due to its effect on aluminum materials. Therefore, we recommended that mercury removal beds be installed to minimize the risk of accumulation of mercury in the system.
(5) To have an LNG product, net in storage, containing 50 ppm or less CO_2, the feed gas to the Liquefaction Unit must contain not more than 40 ppm of CO_2. This allows for 10% boil-off and 5% margin.

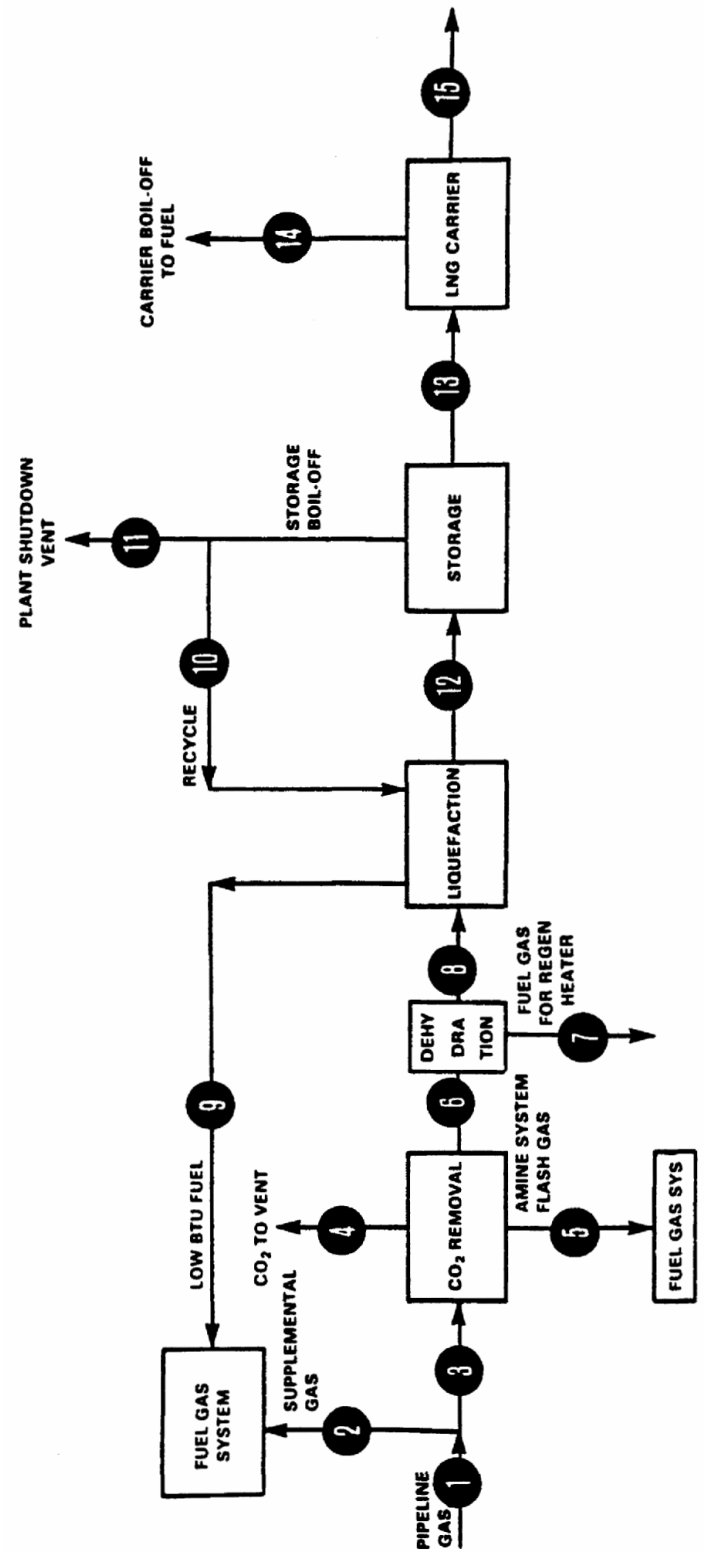


Figure 23.1: Plant Block Flow Diagram

The dehydration is carried out using molecular sieves as the adsorbent. This process has also been described previously in Chapters 5 and 17.

If heavy hydrocarbons, ie. hexanes plus, are present in any measurable amounts, these are taken out by increasing the refrigeration duty and making sure the condensed hydrocarbon is removed from the system prior to entering the liquefaction section.

Mercury, which is sometimes found in small quantities, should be removed to protect the aluminum in heat exchangers. This material is removed using a fixed bed sulphur impregnated carbon adsorption process. This is generally the first step in a gas treating process, if it is needed.

23.5 Liquefaction Processes

There are three main types of liquefaction processes. These are the expansion system, the cascade system, and the mixed refrigerant system. These are discussed further on.

23.5.1 Expander Cycle

Figure 23.2 shows a simplified flow sheet of the expander process. In this particular case, it is assumed that no nitrogen or heavy hydrocarbons have to be separated from the natural gas. The heat of condensation is removed using a closed expander cycle. The circulating refrigerant (nitrogen or nitrogen and methane) is compressed in compressor “C1” and in expander compressor “C2”, which is directly coupled to expansion turbine “ET”.

Compressed refrigerant and natural gas are heated and exchanged in “E1” with expanded refrigerant until the temperature is lowered to the level required at the turbine inlet. Following this, the greater portion of the circulating refrigerant is expanded in the turbine. In exchanger “E2” natural gas is liquefied by heat exchange with the expanded refrigerant. Together with the residual high pressure stream the liquefied gas is then subcooled to the point required for expansion into storage with the minimum flashing.

The Q/T diagram of a process of this kind shows that there exist relatively large temperature difference in the cycle and, because the circulating refrigerant is precooled by heat exchange with itself, also very high heat exchange takes place (about three times the heat of liquefaction). The specific energy requirement is relatively high (about 0.65 kWh/m^3 of LNG at natural gas pressure of 50 atmospheres, as compared with a theoretical energy input of about 0.11 kWh). However, because of its relatively simple operation and reasonable investment costs, this method of liquefaction is particularly well suited to the requirements of small volume peak shaving plants.

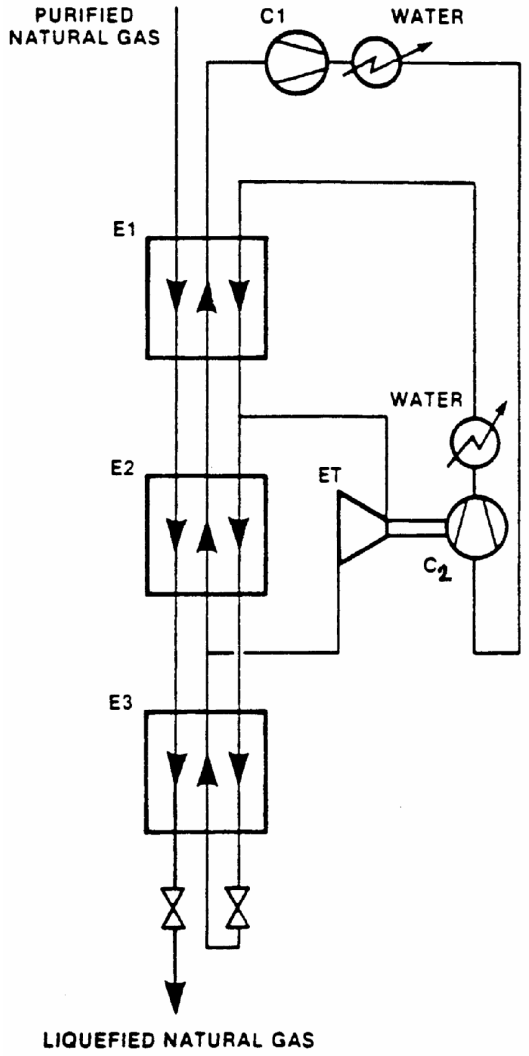


Figure 23.2: Natural Gas Liquefaction Expander Cycle

23.5.2 Cooling Curves

Before discussing the next two processes, ie. the cascade process and the mixed refrigerant process, we should examine a typical LNG cooling curve and how the above mentioned processes follow this curve.

Figure 23.3 shows an LNG cooling curve. Figure 23.4 shows a three level Cascade cooling system (ie. 3 refrigerants - propane, ethylene, and methane). Figure 23.5 shows a nine level Cascade system (ie. 3 refrigerants each with three evaporation temperatures), and Figure 23.6 shows the propane mixed refrigerant process in which the mixed refrigerant is evaporated and the LNG is condensed. Figure 23.7 shows a comparison of the 9 level Cascade and the propane MCR process. The object is to recover as much as possible of the heat removed by heat exchange and also minimize the power required for the process, thus the close approach is the best. Another mixed refrigerant may be substituted for the propane in the process shown in Figure 23.6.

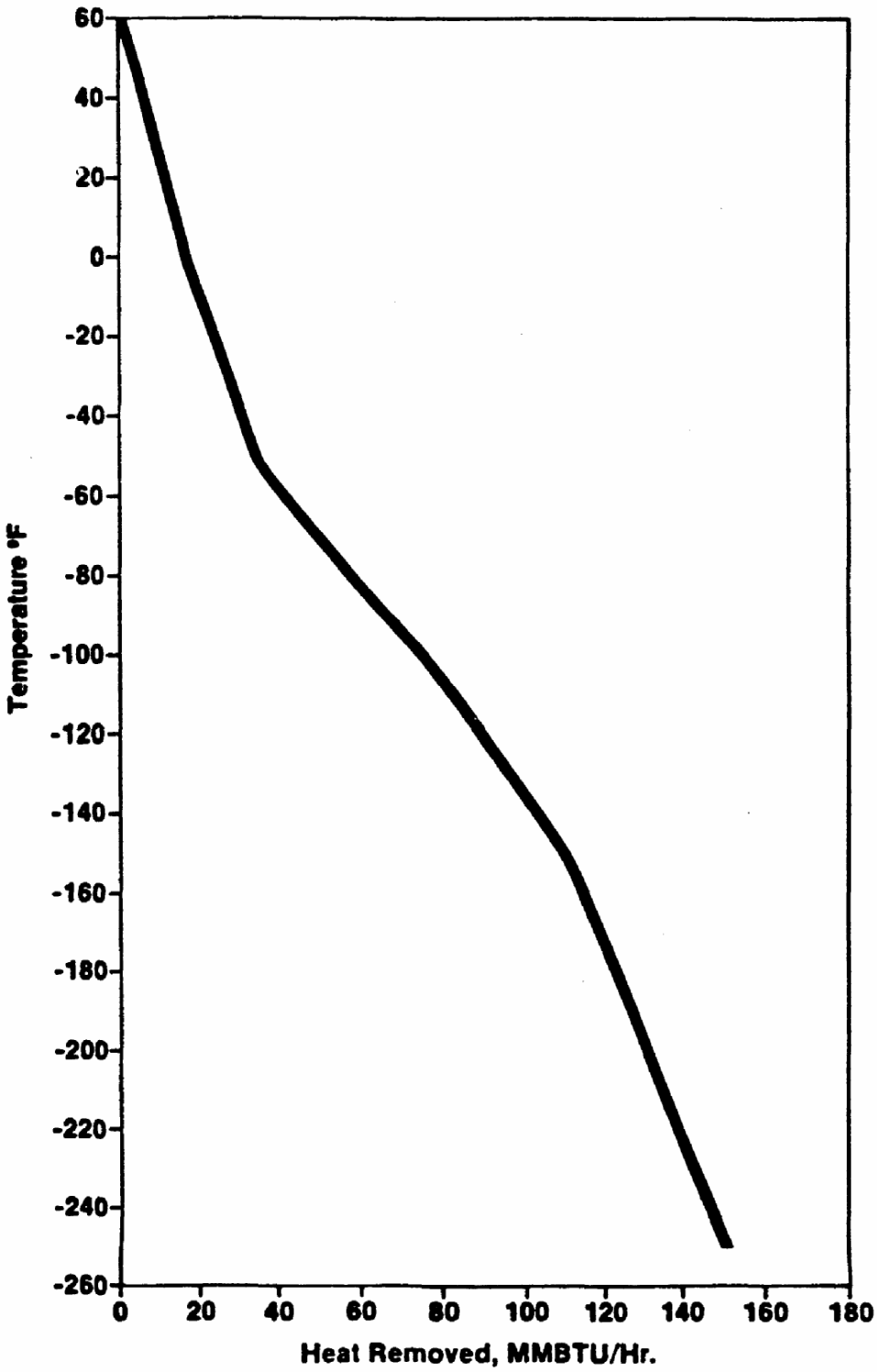


Figure 23.3: Heat Removal as a Function of Temperature for LNG Plant Feed Gas

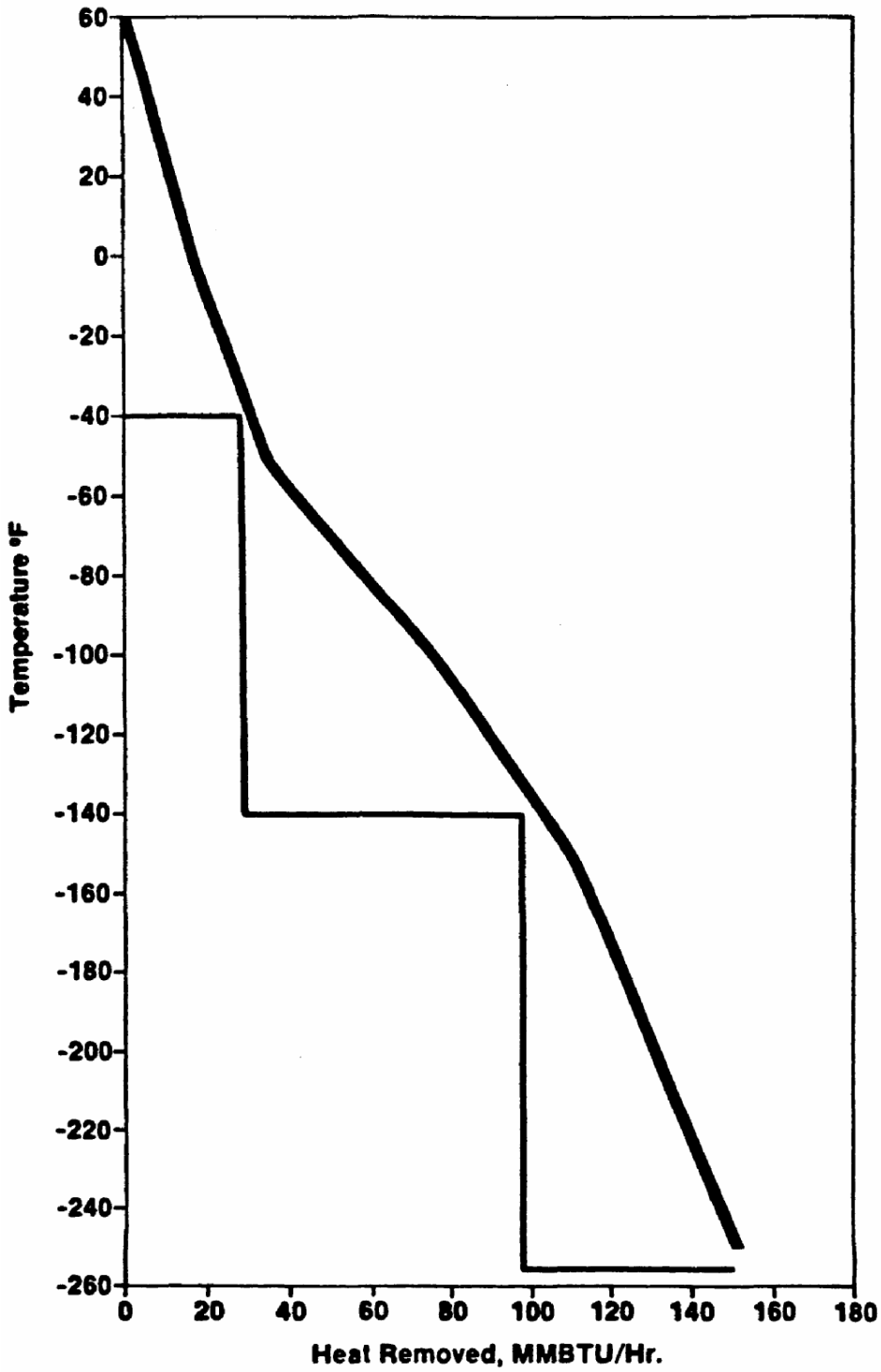


Figure 23.4: Three Level Cascade Cycle Cooling Curve Vs Natural Gas

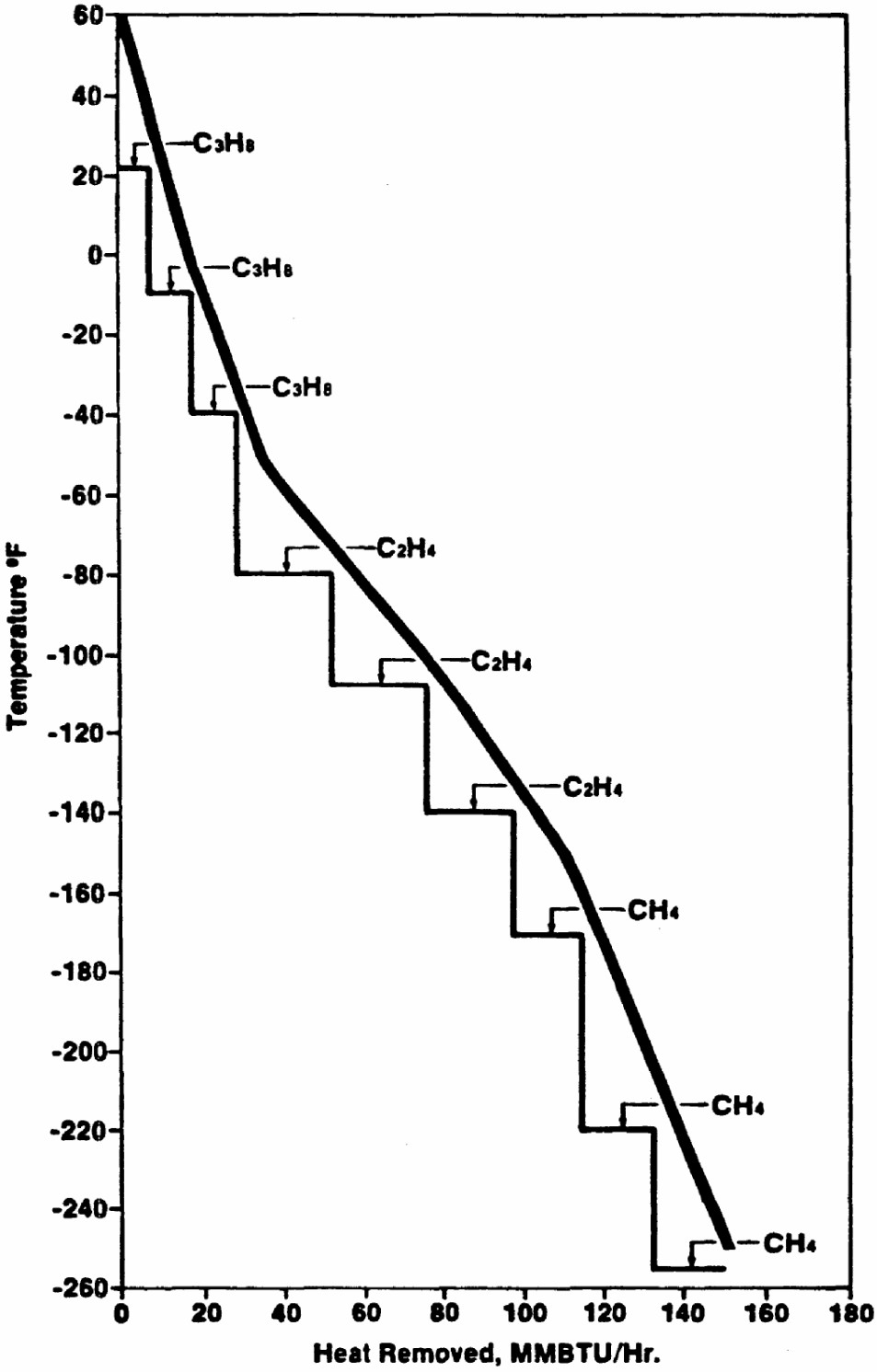


Figure 23.5: Nine Level Cascade Cycle Cooling Curve Vs Natural Gas

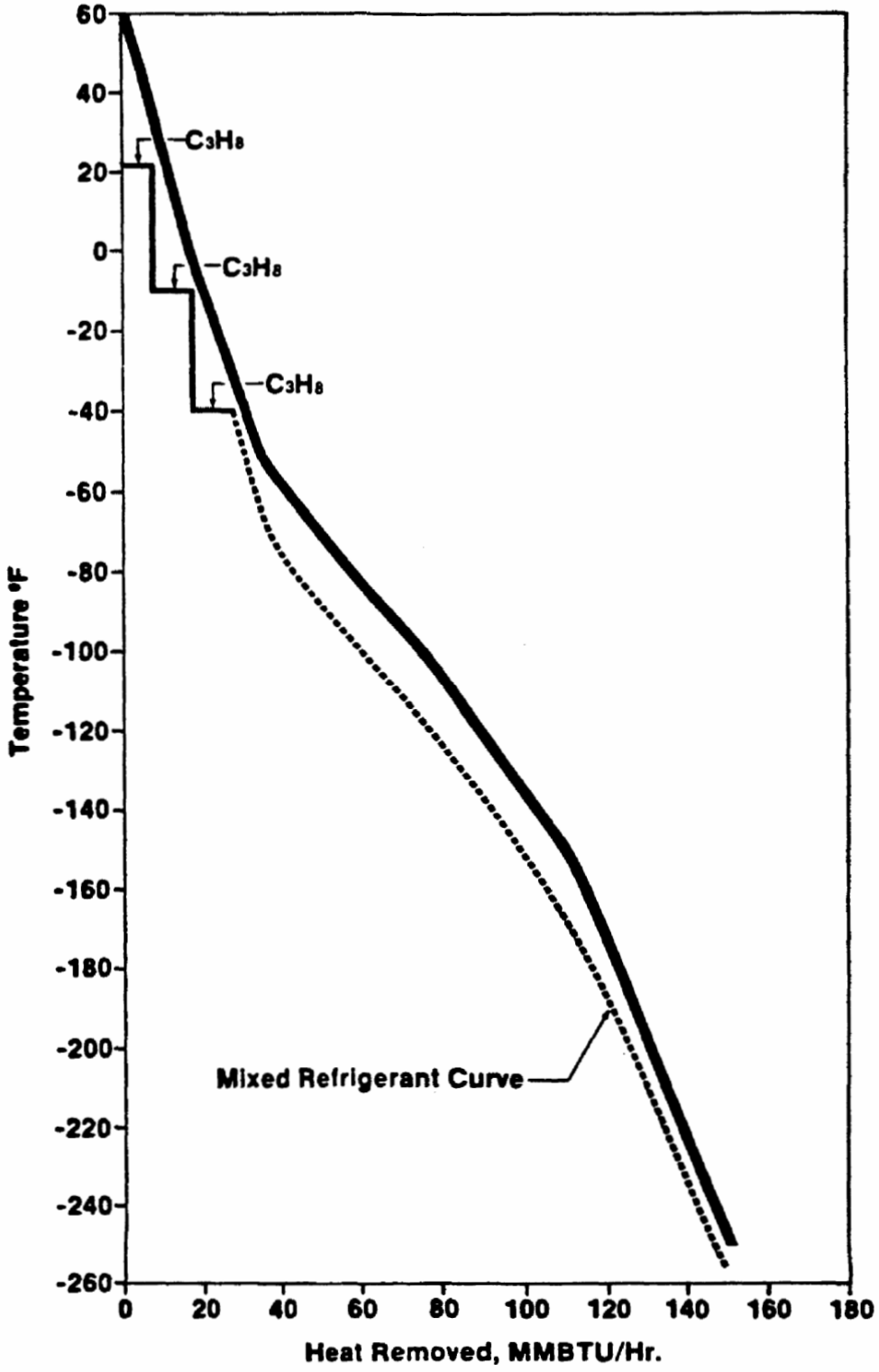


Figure 23.6: Propane Precooled Mixed Refrigerant Cycle Cooling Curve Vs Natural Gas

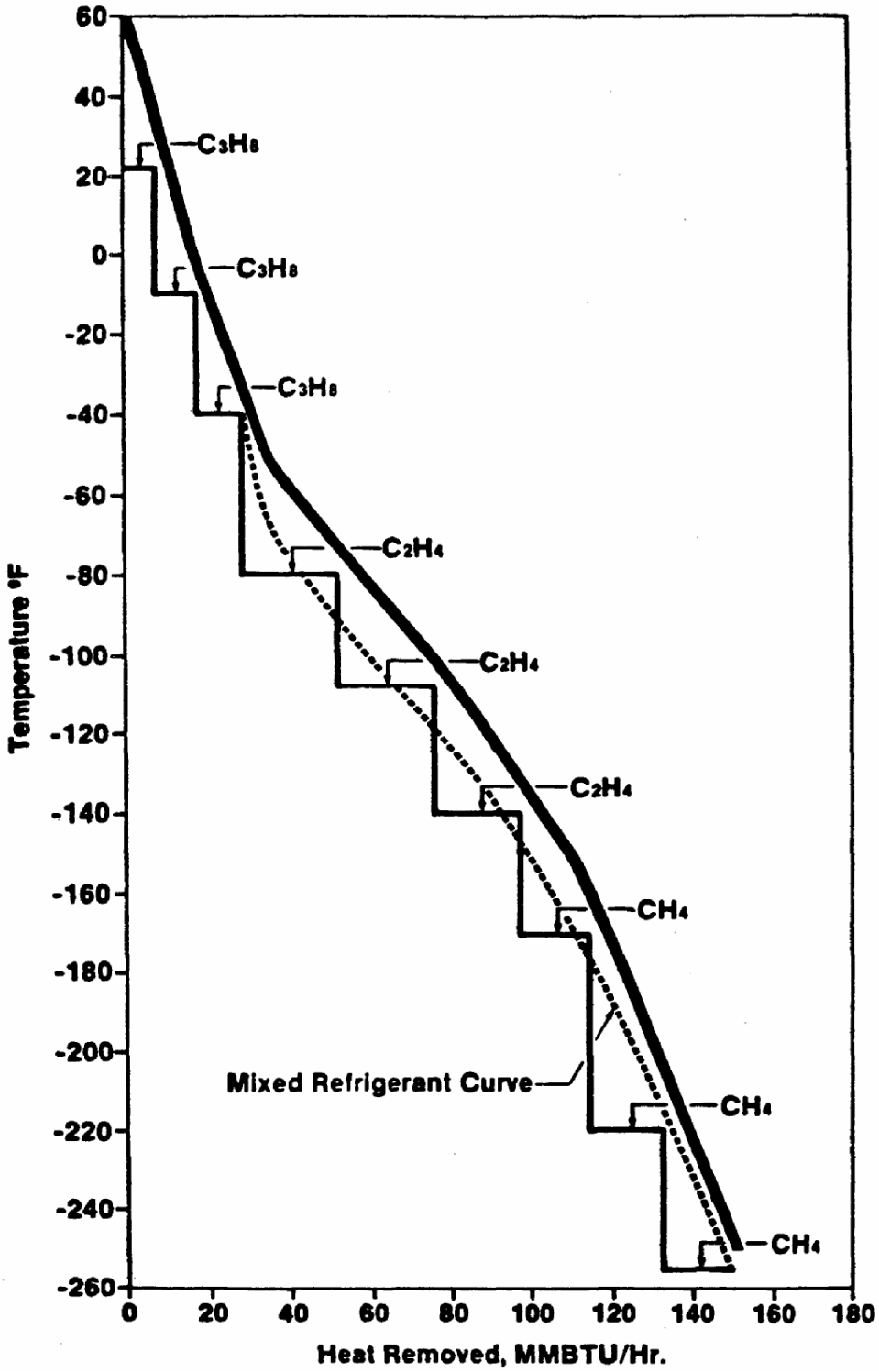


Figure 23.7: Comparison of Nine Level Cascade System Vs Propane Precooled MCR System

23.5.3 Cascade Liquefaction Cycle

The classical Cascade process reduces the irreversible heat exchange losses by utilizing several refrigeration cycles whose refrigerants vaporize at different but constant temperatures. The process flow sheet may be found on Figure 23.8.

Propane is liquefied by heat exchange with cooling water and then expanded in stages to 5, 2.5, and 1 atmosphere with the result that cooling zones at temperatures of about 0ºC, -20ºC, -40ºC respectively are present in the propane cycle vaporizers. In this way, natural gas and cycle gas are cooled in stages. Propane vaporized at the various expansion levels will be cycled to corresponding inlets in the propane recycle compressor.

Ethylene is compressed to about 310 psia and is liquefied in the coldest of the propane evaporators, and is then used as a refrigerant - again in three stages of cooling. In the coldest ethylene evaporator, methane and natural gas are liquefied.

The liquefied methane is used at three different temperature levels as a refrigerant, in the same way as propane and ethylene are used in the early process. Following condensation, the methane for the three cooling stages will be returned to the methane cycle compressors.

Liquefied natural gas taken off in the last and coldest of the methane cycle vaporizers is expanded to tank storage pressure. The specific energy requirements of the classical Cascade liquefaction cycle approaches the upper limit of maximum attainable thermodynamic efficiency; however, the requirement for machinery, heat exchangers, separators and condensers, and control systems is exceptionally high. The first two base load liquefaction plants at Kenai in Alaska and Arzew in Algeria used this liquefaction cycle.

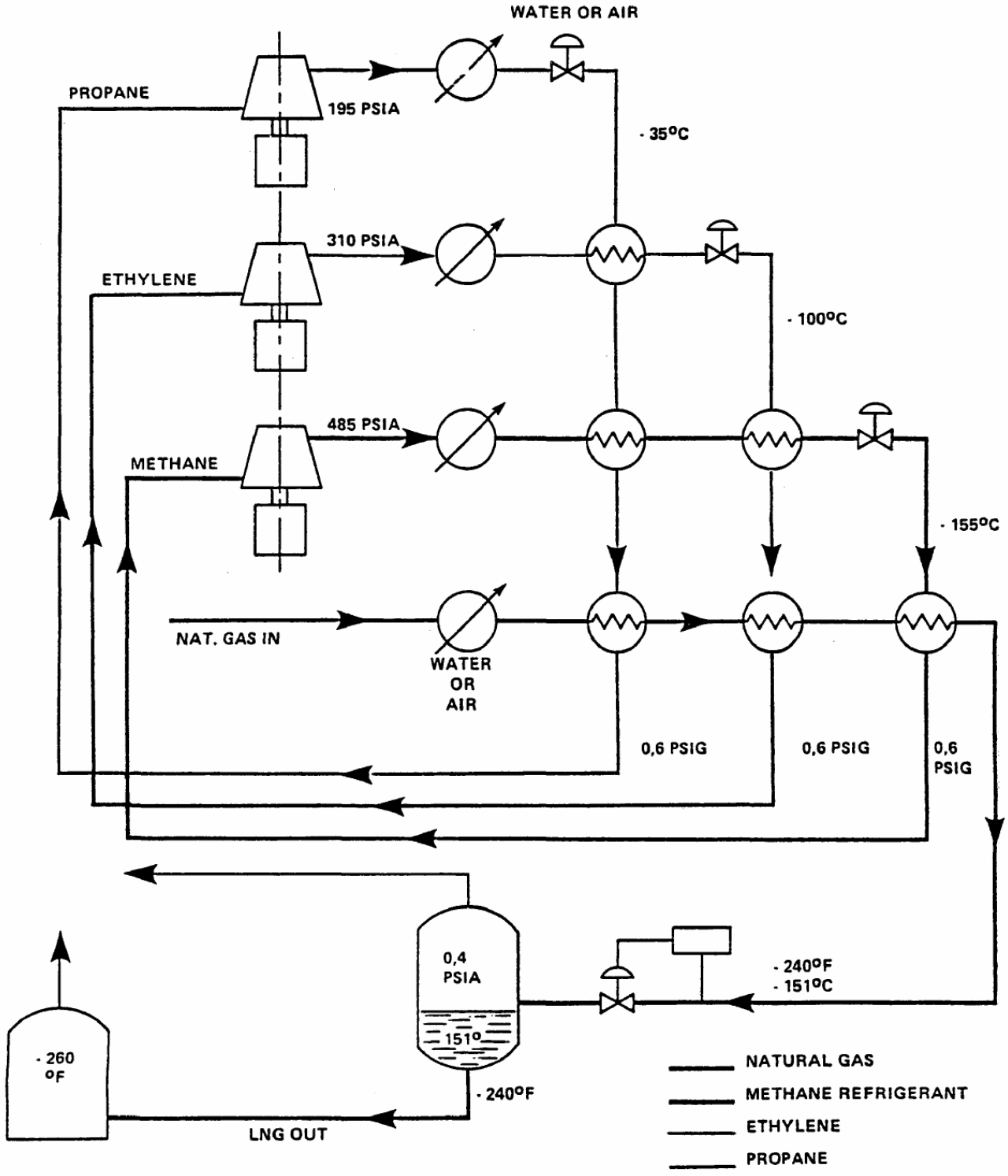


Figure 23.8: Liquefaction of Natural Gas Classical Cascade

23.5.4 Air Products Propane Precooled Mixed Refrigerant Process

The flow sheet for this process may be found on Figure 23.9. The gas stream Is cooled directly as shown. The refrigeration stream flow is more complicated and is as follows:

The low pressure refrigerant stream leaving the cryogenic heat exchanger is compressed to a high pressure in a compressor and cooled to about -35ºC by propane refrigeration in the evaporators. The partially condensed stream is separated into liquid and vapour fractions in a phase separator. The liquid is cooled in the warm bundle of the cryogenic heat exchanger, flashed across a pressure letdown valve, and distributed over the shell side of the exchanger. The vapour stream is condensed and subcooled in both the warm bundle and the cold bundle, flashed across a letdown valve, and distributed over the shell side of the cold bundle. The stream from the bottom of the cold bundle mixes with the low pressure stream before redistribution over the warm bundle,

The propane streams from the high, medium, and low level evaporators of the plant are compressed in a single multi-stage compressor system. The high pressure propane is cooled and condensed against cooling water or air and stored in an accumulator that supplies propane to the evaporators.

The three primary operating parameters for the entire mixed refrigerant system are: composition, suction pressure, and discharge pressure. In the propane system, the low level evaporator pressure is set to obtain the desired temperature of the feed and mixed refrigerant streams to the cryogenic heat exchanger; the high level evaporator pressure is set sufficiently high to avoid hydrate formation in the feed gas; and the medium level evaporator pressure is allowed to float as dictated by compressor performance. The cooling water temperature determines the discharge pressure of the propane compressor. Thus, the only primary control parameter in the propane cycle is compressor speed.

Air Products have reported that the propane precooled mixed refrigerant process can be operated to accommodate large turndown in LNG production. At reduced capacity the efficiency of the mixed refrigerant system can be maintained by changing composition and lowering suction pressure. Similarly, by adjusting the composition of the mixed refrigerant and changing the suction and discharge pressures of the mixed refrigerant compressor, two different feed streams can be liquefied at the same plant without any significant reduction of efficiency. Their conclusion was, that by adjusting the primary operating variables of composition, suction pressure, and discharge pressure, close to optimum efficiency can be maintained even with significant changes in LNG production rates, LNG composition, and off-design equipment performance.

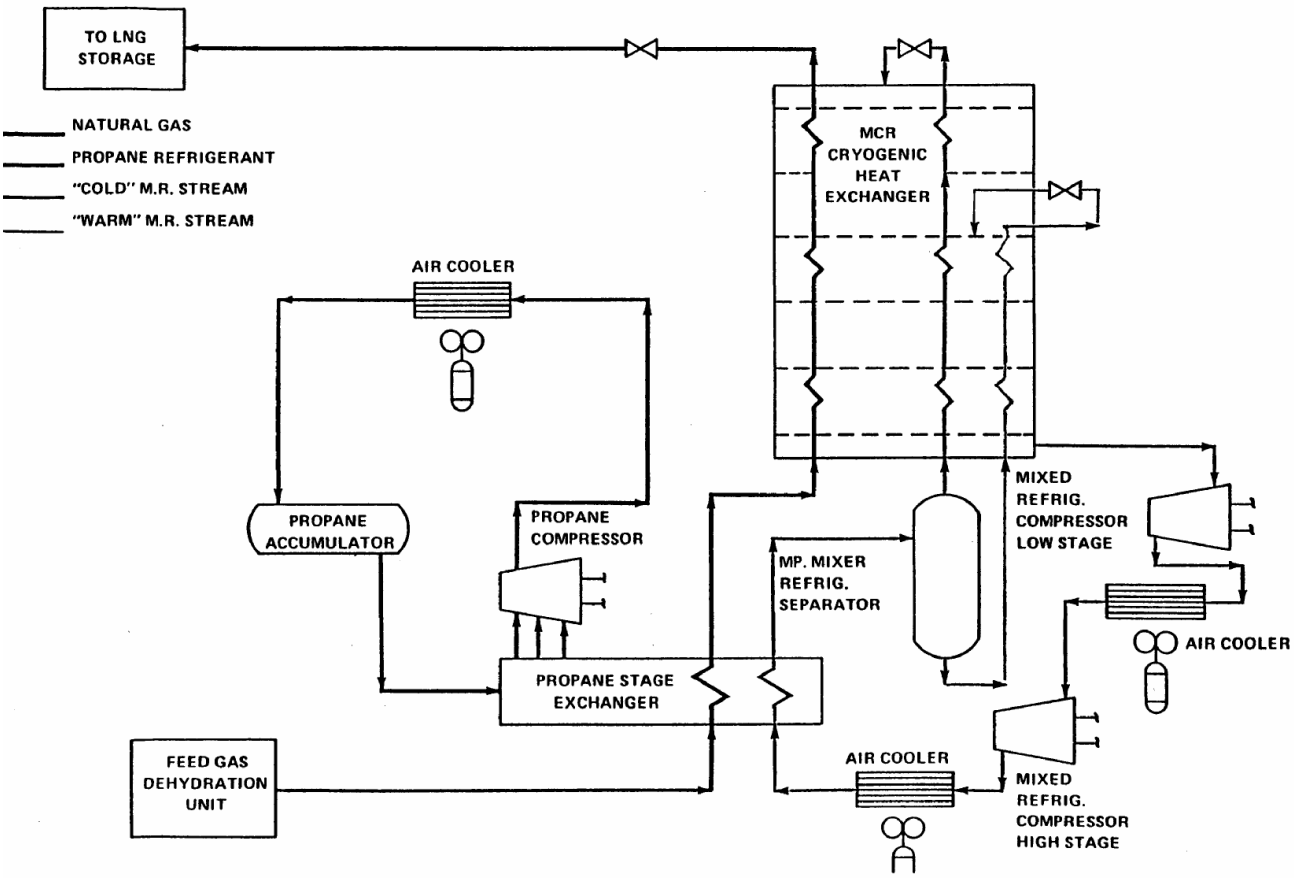


Figure 23.9: Liquifaction of Natural Gas Air Products Propane Precooled Mixed Refrigerant Process

23.5.5 Technip Precooled Tealarc Process

The precooled Tealarc process is basically the same as the propane precooled process supplied by Air Products, with the exception that the precooling refrigerant is a mixture of ethane, propane, and butane rather than pure propane. The flow sheet is on Figure 23.10.

Once again, the refrigeration stream flow is complicated; the low temperature mixed refrigerants, which is a mixture of nitrogen, methane, ethane, and propane, is compressed in two cases of a centrifugal compressor, cooled by water cooling, and then partially condensed in the precooling cycle by the precooling refrigerant. The liquid portion of this mixed refrigerant is condensed in the warm section of the cryogenic heat exchanger, expanded through an expansion valve, and then vaporized into the warm section of the exchanger where it performs its cooling function.

The vapour stream from the separator is also passed through this warn section of the exchanger where it is partially cooled and condensed. It is then further cooled and condensed in the cold section of the cryogenic heat exchanger, then itself is expanded through a Joule/Thompson valve and fed into the top of the cold section of the exchanger. The vaporized mixed refrigerant leaves the bottom of the cryogenic heat exchanger and is cross-exchanged with the incoming natural gas before being fed back to the compressor suction.

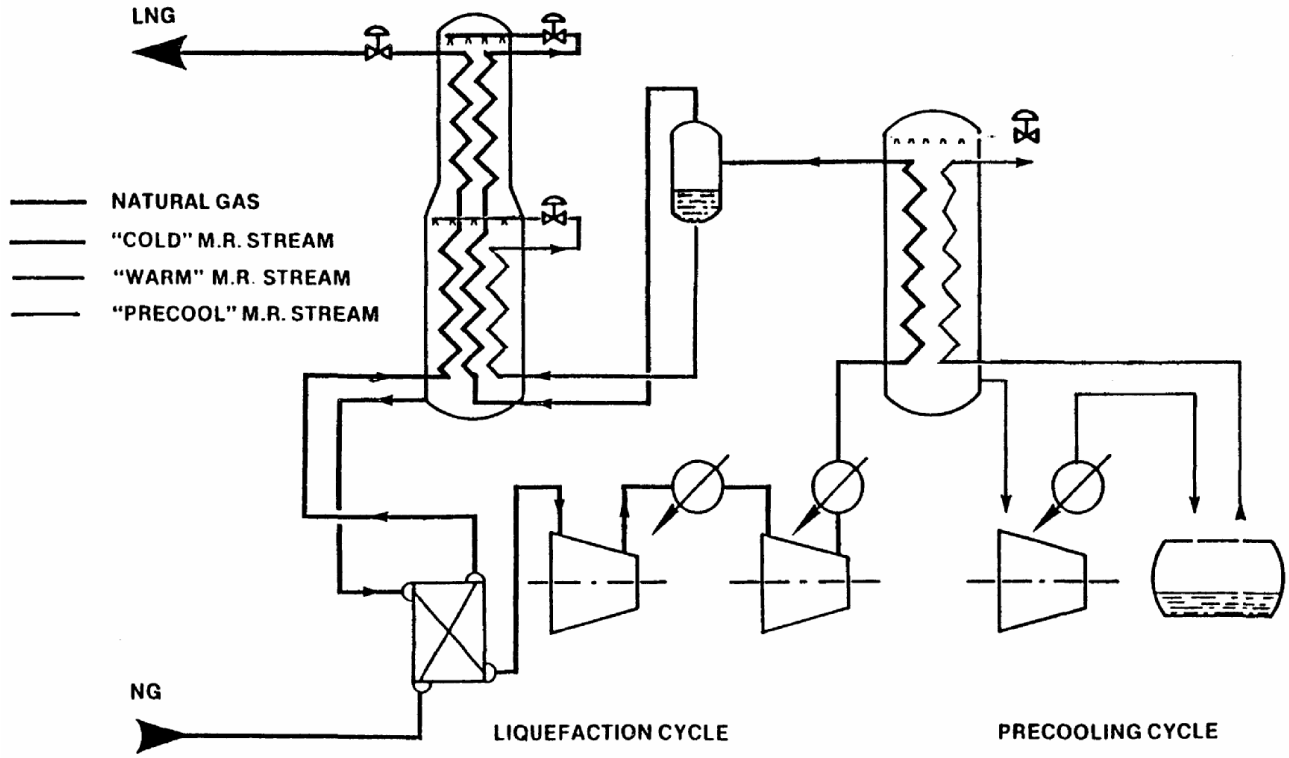


Figure 23.10: Liquifaction of Natural Gas Precooled Tealarc Process

23.5.6 Special Equipment in the Liquefaction Process

There are two types of equipment that are the key to the liquefaction process: the compression equipment and the heat exchange equipment.

23.5.6.1 Compressors & Drivers

Two types of compressors are employed:

- Centrifugal.
- Axial.

The axial type is just starting to be used in existing plants, but is being considered for the mixed refrigerant type plants. Single line units of up to 135,000 hp have been built. The centrifugal type compressors have been used for the existing plants, and are generally being considered for both the Cascade and MR plants now under study. The centrifugals used are of fairly standard, proven design, of large size but for relatively low pressure. MR type plants require a discharge pressure on the order of 450 psig, and for the Cascade type plant, the discharge pressure range about 200 psig for propane, 230 psig for ethylene, and 620 psig for methane.

For the large Cascade type plants, the centrifugal compressor lines being considered range up to 55,000 hp requiring one body for propane, two bodies for methane, and two bodies for ethylene. For MR type plants, the number of compressor bodies vary widely depending upon the manufacturer and the line size. The compressors are provided with intermediate nozzles for entry of economizer streams.

Drivers used have been of two basic types:

- Steam turbines.
- Gas turbines.

Note: Large electric motors have also been considered.

The gas turbines have the advantage of a simpler installation, little water requirement, and usually lower initial cost. Gas turbines are available of proven design up to 64,000 hp. Steam turbines are well proven as variable speed compressor drivers up to 30,000 hp; experience with larger units have been primarily as constant speed electric generator drivers. Steam turbine systems may be slightly more efficient than industrial gas turbines without waste heat recovery, but the efficiency comparison varies for each plant situation. It depends on steam conditions, design of steam system, and the particular gas turbine design, including any gears. Each type of driver has its advantages, and should be evaluated for a specific plant location and situation.

Steam turbines have an efficiency that is almost independent of load while normal overall efficiency is about 31% (on LHV). Also, steam turbines easily fit with compressor power and speed requirements.

Steam, however, has serious disadvantages because in most sites fresh water has to be recovered from sea water; and because it requires large quantities of cooling water.

Two types of gas turbines are available:

- Industrial gas turbines, heavy duty type, where efficiencies reach 27%.
- Jet type turbines derived from aircraft engines in which efficiencies may exceed 30%.

For cold climates, gas turbines are very attractive because they are efficient and do not require cooling water.

The specific fuel consumption of gas turbines decreases at lower temperature such as the following example for the Coberra 6256:

Temp. ºF	BTU/BHP/hr
45	7290
26	7210
-8	7040
-65	6790

This is quite different from the HP output which, for this unit, can increase by 40% between an ambient air temperature of 85ºF and of -8ºF.

Electric motor drives are being planned for the Western LNG project. The selection of electric motor drives is based on the availability of electrical power and its relative cheapness as compared with gas. Also, there has been considerable experience in using large electric motors in ethane extraction plants and gas compression facilities in Alberta (in the 30,000 hp or bigger size).

A thorough review of the appropriate driver and compressor type is most important so as to produce LNG at the minimum cost.

23.5.6.2 Heat Exchangers

Four basic types of heat exchangers have been employed. These are:

- Air fin coolers.
- Conventional shell and tube.

- Brazed plate-fin type.
- Wound type exchangers

The air fin coolers are only suitable for cooling to within 20ºF to 40ºF of the ambient air and are used as compressor inter and after coolers. The shell and tube exchangers are normally used at moderate temperatures, down to approximately -50ºF, as water coolers and precoolers. These have been discussed in Chapter 8. The cryogenic (cold box) exchangers are normally either of the brazed type or wound type. These will be discussed further.

Brazed Plate-Fin Type - These are made from a stack of layers, each one consisting of corrugated aluminum sheets (fins) between flat aluminum separating plates to form individual fluid passages. Each layer is closed at the edge with solid aluminum bars, and the stack is bonded together by a brazing process to yield an integral rigid structure with a series of passages. These passages normally have integral welded headers. Several sections may be connected together to form one larger exchanger.

Advantages:

- Compact; about nine times as much surface per cubic foot as shell and tube exchangers.
- Weight is kept to a minimum.
- Design pressures go up to 1440 psig and from cryogenic temperatures to +150ºF. This type of exchanger is used extensively in ethane extraction plants.

Wound Type - Wound type exchangers are made by spirally winding layers of small aluminum tubes, with spacers. The tubes tie into headers and the shell side (low pressure) fluid flows through the bundle, around the tubes. The tube side may be provided with headers to allow multiple separate flows and functions.

Advantages:

- They can be made huge with multiple bundles in one shell, eliminating some headers, piping, etc.
- They can be built for higher tube side pressures, up to about 2,000 psig.
- Spiraling of tubing eliminates contraction problems. The article “The Design Fabrication and Large Cryogenic Heat Exchangers” by L.S. Gaumer et al discusses these exchangers in some detail.

  The maximum present size of this equipment is about 15 feet (4.6m) in diameter, 250 feet (76m) long, and weighs about 400 tons (367 tonnes). This size is restricted primarily by transportation restrictions.

23.5.7 Final Flash or Sub-cooling

LNG is stared at 15 psia and has to be produced in the liquefaction unit at a temperature close to the bubble point at storage pressure (about -260ºF).

This can be achieved in two ways:

- LNG sub-cooling.
- LNG final flash.

In the LNG sub-cooling process, LNG is produced at -260ºF and flows to storage under flow control.

In the final flash process LNG exits from the cryogenic exchanger at -230ºF.

The final decrease of temperature is achieved by partial vaporization of the LNG at 15 psia. The remaining LNG is pumped to storage while the vapour is heated and compressed to fuel gas pressure.

Additional items of equipment are required for the final flash scheme:

- Flash drum.
- LNG pump.
- Flash gas heater.
- Flash gas compressor.
- Flash gas after-cooler.

However, the theoretical work of liquefaction is 0.9% lower per % of flash in the flash drum. Normally, the flash drum volume is 10%, thus, the power reduction is 9%. However, additional power is required to increase the flash gas pressure to fuel gas pressure - 60 psig for a steam plant and 250 to 450 psig for gas turbine plant.

Final flash has an advantage in that it achieves a selective denitrification of the LNG up to 8% N_2 in the inlet gas.

23.5.8 Comparison of LNG Processes

The various processes can be compared on a number of points:

- The number of equipment items required.
- Compressor horsepower required.
- Simplicity of control and turndown.
- Proven capability.
- Area of heat exchanger.

On the number of equipment items, the Cascade type of plant has approximately 50% more equipment items than the mixed refrigerant processes. This in turn leads to a degree of complexity and difficulty in control and turndown and will result in a plant cost of about 20% greater than the mixed refrigerant processes.

At one time it was argued that the Cascade process was considerably more efficient than the mixed refrigerant processes, but, with the development of the propane precooled and the precooled Tealarc, mixed refrigerant processes have achieved efficiencies in terms of compressor horsepowers equal or better than the optimized Cascade process.

It should be noted that all, base load LNG plants built since 1969 (see Table 23.1) have been the mixed refrigerant process.

A close review of the compressor power required, the heat exchanger surface proposed, and the closeness of the temperature approach in the main liquification heat exchange equipment should be made: particularly the actual design calculations and guarantees should be reviewed.

The compressor power basically sets the energy required by the process although other items also affect the total energy consumption as noted later. The energy consumption is the major operating cost of an LNG plant.

23.5.9 Power Required for the Process

Work Required

A formula for the calculation of the power absorbed by the refrigeration process is given:

Actual Work for NG Liquefaction = Theoretical Work for NG Liquefaction + Heat Rejection Temperature x Irreversibilities due to Exchanger Approaches + Refrigeration Process Irreversibilities due to Friction Loss

Equation 23-1

$$W_2 = W_1 + T_2\,\Delta S + \Delta W_3$$

Equation 23-2

$$W_1 = Q\frac{(T_2 - T_1)}{T_1}$$

where:

W_2 = total work required for liquefaction.

W_1 = theoretical work required. Generally you start with natural gas in a gaseous state at room temperature and high pressures and finally store it as a liquid at -260ºF (-162ºC) and atmospheric pressures.

T_2 = the heat rejection temperature in degrees absolute. It is affected by climatic conditions and in the Arctic results in a 25% reduction in power requirements.

T_1 = the liquefaction temperature in absolute.

Q = the enthalpy difference between the gas entering the facility and liquid leaving same.

ΔS = the refrigeration process irreversibilities. It is affected by the refrigeration process used and how well the refrigerant Q vs T curve matches the liquefaction Q vs T curve.

ΔW_3 = the work required to overcome friction is affected by pressure drop in the exchangers and piping.

The Figure 23.11 shows the theoretical work required to liquefy a lb of methane from a heat rejection temperature 105ºF to 258ºF and 14.7 psia and the expected extra work if a 4 stage machine is used to carry out this equation. The closer the refrigeration evaporation curve can approach the cooling curve for the methane the less extra power is required for this operation.

The effect of pressure of liquefaction and of the composition of the feed stock has been studied. This data is given in table that follows the discussion below.

The results showed that, to minimize HP/MMSCFD of LNG, the liquefaction has to be performed at high pressure: a difference of 12% occurs between 500 psia and 800 psia.

Note: If the feed gas contains heavy hydrocarbons, a separation of NGL is required, and this limits the liquefaction pressure to 700 psia and perhaps lower.

When the inlet gas is at low pressure it is worthwhile to increase its pressure above 700 psia in order to reduce the total HP required. For a dry gas at 440 psia it was determined that the optimum pressure was 1000 psia.

For a specific LNG unit operating at constant compression power, if the operating pressure decreases, the LNG production also decreases due to the increase of theoretical work of liquefaction.

A study of the influence of composition was made for four feeds, ie., a heavy gas, a medium gas, pure methane, and a nitrogen-methane mixture. The results were:

	BHP/MMSCFD	%
Heavy Gas	393	95
Medium Gas	413	100
Pure Methane	420	102
Nitrogen-Methane	447	108

If the unit is to process different feeds, the LNG production may be increased when processing a heavy feed. This may be done with a flexible process.

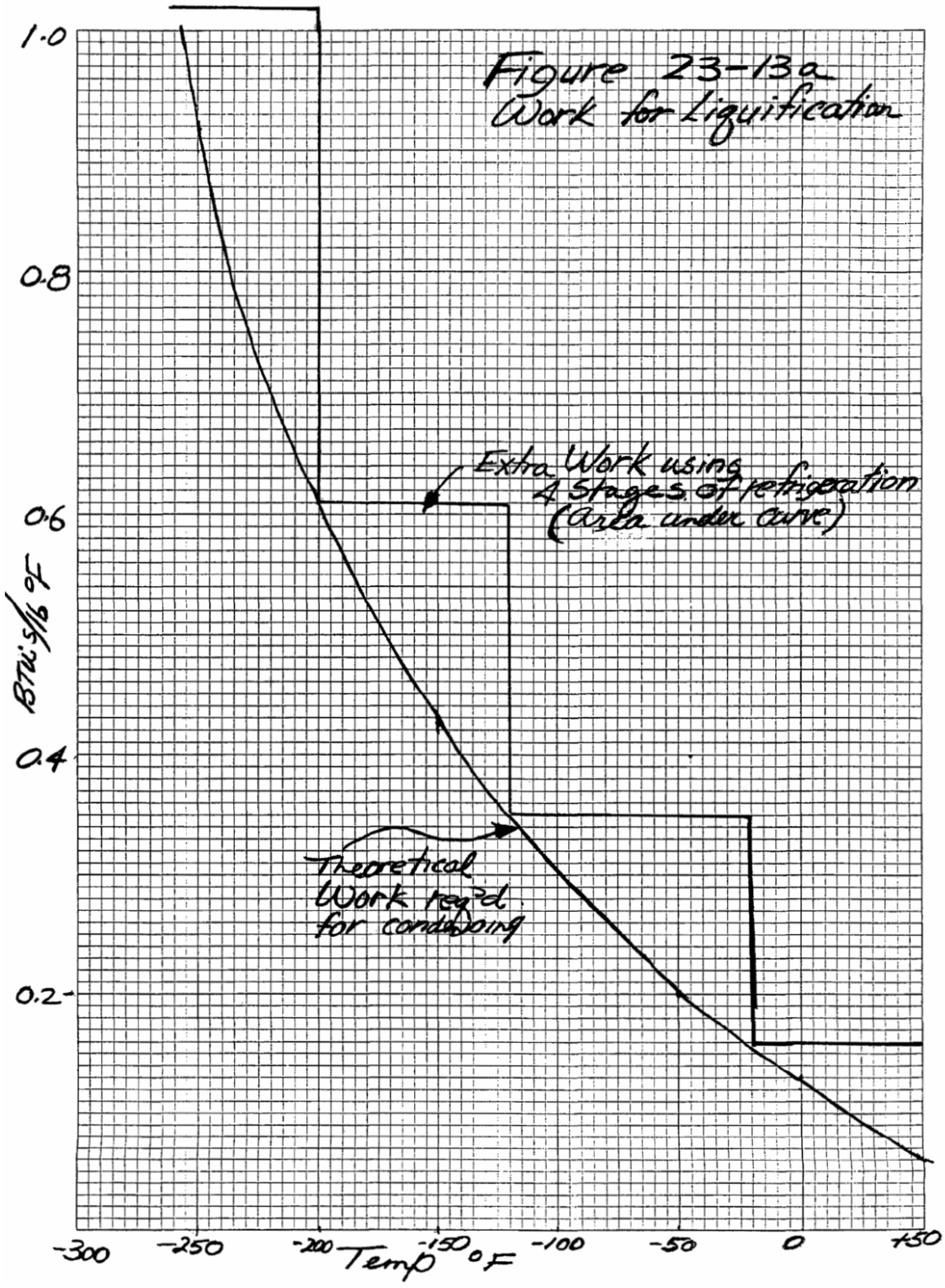


Figure 23.11: Work for Liquefaction

23.5.10 Optimization of Work Required

From an energy point of view refrigeration involves two steps:

- Transformation of the liquefaction work into work to be performed on the refrigerant, which is achieved in the cryogenic heat exchanger.
- Compression and cooling of the refrigerant by absorption of work and rejection of heat.

Consequently, to optimize efficiency of the cycle both the exchanger and the compressor efficiency have to be simultaneously optimized.

To increase the cryogenic exchanger efficiency two things are important:

- reduction of exchanger approach ΔT. This is especially the case at low temperatures. This leads to an increased exchanger surface.
- reduction of heat exchanged. This is done by reducing the total weight flow by using a high compression ratio. The reduced refrigerant flows to the cold exchangers is done by separation of the refrigerant into several fractions used at different temperatures.

Irreversibilities in the compression system are mainly due to the fact that the compression is not isothermal. Note that high compressor efficiencies are obtained with low compression ratio while the opposite is true for the cryogenic heat exchanger. In practice, the optimum compression ratio is between 3 and 6 per stage depending on the k value of the gas.

Of the two types of compressors available the axial flow compressor can have a polytropic efficiency of 85% while centrifugal compressors only reach 78%.

Best overall efficiencies are obtained in processes using two cycles such as propane precooled MR (Air Products) and MR precooled MR (Linde and Technip). In these processes a single refrigerant is used to achieve liquefaction - it is a mixture of nitrogen, methane, ethane or ethylene, and propane. This MR is precooled either with a propane cycle or with another MR which, in one case, is a mixture of ethane, propane, and butane.

Total Energy Consumption

For any LNG process eight important factors affect energy consumption:

Process Selection Factors

- Final Flash or Subcooling.
- Refrigeration Process.
- Mechanical Power System.
- Cooling System.

Design Parameters

- Inlet Pressure and Temperature
- Gas Composition
- NGL Extraction (Composition Related)
- Climatic Condition

23.6 Storage

A very important part of LNG plant and receiving terminals is the storage facilities. Table 23.1 shows the type of storage used in the base load LNG plants, Table 23.3 shows the types of storage used in receiving terminals.

The type of LNG storage for a particular installation should be selected based on safety, reliability, cost, and boil-off rate, as prime considerations. There are the following two main types of LNG storage in general use; both generally store LNG at atmospheric pressure and low temperature:

- Conventional above-ground, double walled metal tanks.
- Prestressed concrete tanks, above or below ground.

As can be seen by the Tables, the majority of the storage tanks are the conventional above-ground, double walled metal tanks.

Other LNG storage of less importance are in-ground frozen earth storage and mined caverns. In-ground frozen earth storage has been utilized; however, high boil-off rates and certain unknowns and disadvantages have restricted its use. Mined caverns are still in the experimental stage for LNG.

Table 23.3: Import Receiving Terminals - Japan

Company and Site	Storage Capacity, 10^6 CF	Storage Capacity, 1000 kl	Capacity	Regasification-10^6 CF/day Number	Regasification-10^6 CF/day Type	Type of Storage Container	Storage Contractor	System Contractor	Year of Operation
TOKYO GAS CO./TOKYO ELECTRIC Ogishi, Yokohama, Japan	3460	1010 (3 x 204) (3 x 221)	190 49	6 at 66 1 at 66	Submerged Submerged	Aboveground, 9% nickel	IHI	Tokyo Gas Co.	1969
	215	63	49	1 at 66	Submerged	Inground concrete and SUS membrane	IHI		1970
	1287	376				Inground concrete and SUS membrane	IHI		1972
	979	284	191	1 at 93 1 at 66 1 at 66	Submerged Submerged Submerged	Aboveground, 9% nickel	IHI	Tokyo Gas Co.	1972
	544	158				Aboveground, 9% nickel	IHI		1973
	2038	599	110	1 at 110	Submerged	Inground concrete and SUS membrane	IHI/Kajima	Tokyo Gas Co.	1977
	2038	599	110	1 at 110	Submerged	Inground concrete and SUS membrane	IHI/Kajima		1978
	2038	599	110	1 at 110	Submerged	Inground concrete and SUS membrane	Nippon Kokan	Tokyo Gas Co.	1979 1981
	2038	599				Inground concrete and SUS membrane	Kawasaki Heavy Industries		1982
TOKYO GAS CO./TOKYO ELECTRIC, Sodegaura, Chiba, Japan	1931	546 (2 x 284)	330	3 at 66 2 at 66	Submerged Submerged	Aboveground, Al	IHI	Tokyo Gas Co.	1973
	1287	370	410	3 at 66	Submerged	Aboveground, Al	IHI	Tokyo Gas Co.	1974
	2574	736 (2 x 370)		2 at 110	Submerged	Inground concrete and SUS membrane	IHI-Shimizu		
	1931	546 (2 x 284)				Aboveground, Al	IHI		1974
	2574	736 (2 x 370)				Aboveground, Al	IHI	Tokyo Gas Co.	1975
	1287	370	352	2 at 110 1 at 66 1 at 66	Submerged Submerged Submerged	Inground concrete and SUS membrane	IHI-Shimizu	Tokyo Gas Co.	1976
	2574	736 (2 x 370)				Aboveground, Al	IHI		1977
	5148	1512 (4 x 378)	613	2 at 110 2 at 131 1 at 131	Submerged Submerged Submerged	Inground concrete and SUS membrane	IHI-Shimizu	Tokyo Gas Co.	1977
	2574	736 (2 x 378)				Inground concrete and SUS membrane	Kajima-MHI		1978
	2574	736 (2 x 378)	241	1 at 131 1 at 131		Inground concrete and SUS membrane	IHI-Shimizu MHI-Kajima Kajima-MHI		1979
	2574	736 (2 x 378)				Inground concrete and SUS membrane			1979
	1287	378				Inground concrete and SUS membrane	MHI-Kajima		1981
	8376	2460 (3 x 820)	482	2 at 110 2 at 131		Inground concrete and SUS membrane	MHI-Kajima IHI-Shimizu MHI-Chiyoda		1982
	3868	1136 (2 x 568)				Inground concrete and SUS membrane	IHI-Shimizu IHI-Shimizu/Nasuma		1982
	3868	1136 (2 x 568)				Inground concrete and SUS membrane	MHI-Kajima IHI-Kajima		1982

Company and Site	Storage 10^6 CF	Storage Capacity, 1000 bbl	Capacity	Regasification-10^6 CF/d Rate	Regasification-10^6 CF/d Type	Type of Storage Container	Storage Contractor	Systems Contractor	Year of Operation
OSAKA GAS CO. Senboku I, Osaka, Japan	2901	852 (3 x 284)	160	3 at 40 1 at 40	Seawater Intermed. Fluid	Aboveground, aluminum (1) 9% nickel (2) Ingrond concrete and SUS membrane	IHI	Osaka Gas	1971
	967	284					IHI/Obayashi		
OSAKA GAS CO. Senboku II, Osaka, Japan	16036	4710 (10 x 471)	340	7 at 60 2 at 60	Seawater Intermd. Fluid	Aboveground, aluminum (3) 9% nickel (7)	IHI and Toyo Kanetsu K.K.	Osaka Gas	1977
	4810	1413 (3 x 471)	300	2 at 150	Intermd. Fluid	Aboveground, 9% nickel	IHI and Toyo Kanetsu K.K.	Osaka Gas	1982
	3207	942 (2 x 471)				Aboveground, 9% nickel	IHI and Toyo Kanetsu K.K.	Osaka Gas	U.C.
KITA KYUSHU LNG CO. Yahata, Fukuoka, Japan (Kyushu Electric/Nippon Steel)	7722	2268 (6 x 378)	678	6 at 142 1 at 110	Seawater Submrgd	Aboveground, 9% nickel	IHI and Toyo Kanetsu K.K.	IHI	1977
			142	1 at 142	Seawater				
CHUBU ELECTRIC/TOHO GAS CO., Chita, Aichi Japan	6428	1888 (4 x 472)	668	5 at 142 1 at 66 2 at 22 1 at 72 1 at 59 1 at 66 1 at 44 1 at 7	Seawater Submrgd Seawater Submrgd Seawater Seawater Seawater/ Intermd. Fluid Seawater/ Intermd. Fluid	Aboveground, aluminum (2)	IHI	IHI	1977
						Aboveground, 9% nickel (2)	Toyo Kanetsu		
CHITA LNG CO., New Chita, Aichi, Japan (Chubu Electric/Toho Gas)	6850	2012 (4 x 503)	284	1 at 142 1 at 142	Seawater Seawater/ Intermd. Fluid	Agoveground, 9% nickel (4)	Toyo Kanetsu and IHI	IHI	1983 (U.C.)
	3425	1006 (2 x 503)	284	1 at 142 1 at 142	Seawater Seawater/ Intermd. Fluid	Agoveground, 9% nickel (2)	MHI		1984 (U.C.)
KANSAI ELECTRIC/OSAKA GAS, Himeji, Hyogo, Japan	5138	1509 (3 x 503)	625	5 at 125	Seawater	Aboveground, 9% nickel (1) Aluminum (2) Aboveground, aluminum	IHI	IHI	1979
	858	252					IHI	IHI	1979
	5138	1509 (3 x 503)				Aboveground, 9% nickel	IHI	IHI	U.C.
	6851	2012 (4 x 503)	120	2 at 60	Intermd. Fluid	Aboveground, 9% nickel	IHI, Toyo Kanetsu K.K. IHI	Osaka Gas	U.C.
HIROSHAI LNG Higashie, Hiigata, Japan	6851	2012 (4 x 503)	1250	3 at 193 2 at 198 2 at 39 1 at 131 1 at 66	Seawater Seawater Seawater Submrgd Submrgd	Aboveground, aluminum (2) Aboveground, 9% nickel (2)	IHI MHI	Tokyo Gas Co.-MHI	1983
	4284	1258 (2 x 629)				Aboveground, 9% nickel (1) Aboveground, aluminum (2)	MHI IHI		1984
TOKYO ELECTRIC Higashi-Ogishima, Kawasaki, Japan	2574	756 (2 x 378)	440	3 at 110 1 at 110	Seawater Submrgd	Ingrond concrete and SUS membrane	IHI-Shimizu	Tokyo Electric	1983
	3861	1134 (3 x 378)				Ingrond concrete and SUS membrane	MHI-Taisei		
	2574	756 (3 x 378)				Ingrond concrete and SUS membrane	MHI-Obayashi		1984 (U.C.)

Table 23.4: Import Receiving Terminals - Europe

Company and Site	Storage 10^6 CF	Capacity, 1000 bbl	Capacity	Regasification-10^6 CF/d Runs	Type	Type of Storage Container	Storage Contractor	Systems Contractor	Year of Operation
BRITISH GAS CORP. Canvey Island, England	1400	392 (6-6042-16)		3 at 50 1 at 50 2 at 50	Submrgd Seawater Intermediate fluid	Aboveground, aluminum	APV and Wheeoe	Gas Corp.	1958, 1963
GAZ DE FRANCE Montoir de Bretagne	5148	1512 (2 x 756)	880	4 at 220	Estuary water	Aboveground, concrete	Chantiers Modernes; Europe Etudes; Technigaz; Bignier-Schmide-Laurent	Gaz de France	1981
			300	4 at 75	Submrgd				
	2574	756	300	4 at 75	Submrgd	Aboveground, 9% nickel	CMP	Gaz de France	1983 (U.C.)
GAZ DE FRANCE Le Havre, France	781	229.5 (3 x 76.5)	178	2 at 89	Seawater	Aboveground, 9% nickel	CMP	Gaz de France	1963
GAZ DE FRANCE Fos sur Mer, France	1532	450 (2 x 225)	223	223	Intermed Fluid	Aboveground, aluminum	CMP	Gaz de France	1971
	1712	503	300	6 at 50	Submrgd	Aboveground, 9% nickel	CMP		1979
ENAGAS, S.A., Barcelona, Spain	1800	500 (2 x 250)	155	1 at 67 2 at 44	Seawater	Aboveground, concrete	Preload/Christiani and Nielsen	Esso R & D	1969
	1800	500	268	2 at 67	Seawater	Aboveground, 9% nickel	CMP	Lummus/Gas Natural	1974
	1800	500	536	2 at 67 4 at 134	Submrgd Seawater	Aboveground, concrete	Preload Auxini	Procon/Initec	1981
SNAM (ENI) La Spezia, Italy	2100	630 (2 x 315)	353	2 at 1765	Intermed. Fluid	Aboveground, 9% nickel	SAIPEM (PDM llc.)	Esso R & D	1969
DEUTSCHE FLUSSIGERDGAS TERMINAL, Gmbh Wilhelmshaven, Germany	5300	1508 (4 x 377)	2160	16 at 135	Submrgd/ seawater	Aboveground, concrete and 9% nickel	Dykerhoff Widmann-CMP	Linde A.G.	1980's (planned)
DISTRIGAZ, S.A. Zeebrugge, Belgium	7450	2188 (4 x 547)	800	80 at 100	Submrgd	Half underground, 9% nickel	Frabeco-CPE	Tractebel	1986 (U.C.)

Table 23.5: Import Receiving Terminals – U.S. and Canada

IMPORT/RECEIVING TERMINALS — U.S. AND CANADA

Company and Site	Storage 10^6 CF	Storage Capacity, 1000 bbl	Capacity	Regasification–10^6 CF/d Runs	Regasification–10^6 CF/d Type	Type of Storage Container	Storage Contractor	Systems Contractor	Year of Operation
ALGONQUIN LNG INC. Providence, R.I.	2000	600	300	3 at 100	Direct Fluid	Aboveground, 9% nickel	CBI	CBI	1973
COLUMBIA LNG CORP. & CONSOLIDATED SYSTEM LNG CO., Cove Point, Maryland	5000	1500 (4 x 375)	1000 200	10 at 100	Submrgd Intermed. Fluid	Aboveground, aluminum	PDM	Pullman Kellogg Co. and Raymond Technical	1977
DISTRIGAS OF MASSACHUSETTS CORP., Everett, Mass.	3230	974 (1 x 374 + 1 x 600)	135 150	3 at 45 3 at 50	Submrgd Hot Water Exchangers	Aboveground, 9% nickel	CBI	CBI	1971 1979
SOUTHERN ENERGY CO. Elba Island, Ga.	4000	1200 (3 x 400)	540	5 at 108	Submrgd	Aboveground, aluminum	CBI	Bechtel Inc. and Raymond Technical	1977
TRUNKLINE LNG CO. Lake Charles, La.	6000	1800 (3 x 600)	700	7 at 100	Submrgd	Aboveground, 9% nickel	PDM	Pullman Kellogg Co. and Raymond Technical	1982
WESTERN LNG TERMINAL CO., Point Conception, Calif.	5775	1650 (3 x 550)	1600	13 at 100	Direct Seawater	Aboveground, 9% nickel	N.A.	Fluor	1980's (On hold)

N.A. - Not available.

23.6.1 Above-Ground Double Walled Metal Tanks

Double walled metal tanks are normally flat bottomed with a domed roof, having insulation between the two walls. This allows the outer tank to be of materials suitable for ambient temperatures. The outer bottom must be maintained at a moderate temperature, by either heating or placing on a raised slab so air can freely circulate under the slab, to prevent over chilling and contraction of outer bottom. This also prevents any possible heaving problems.

The inner tank must be of materials that are ductile and tough at the low operating temperatures. Suitable materials are aluminum, austenitic stainless steel, and 9% nickel steel. Tanks are in service with both an inner roof and nitrogen purged insulation space and with a layer of insulation hung from the outer roof and warm LNG vapour in the insulation space. The insulation used between the bottoms must have adequate strength to support the inner tank, while between the vertical walls provision must be made to prevent compaction and high forces. These tanks have proven very reliable, predictable as to boil-off, are highly resistant to fire, and do not require a certain set of geological conditions. Adequate area for diking is required. The dyke space should be adequate to handle a full tank volume. Sizes above 500,000 BBL have been built and are in service. Figure 23.12 shows the Cross Section of a typical double walled tank.

It should be mentioned that most new designs do not have any side penetrations. All in and out connections are through the roof. Figure 23.13 and Figure 23.14 give more details of the tank with either nickel or aluminum lining.

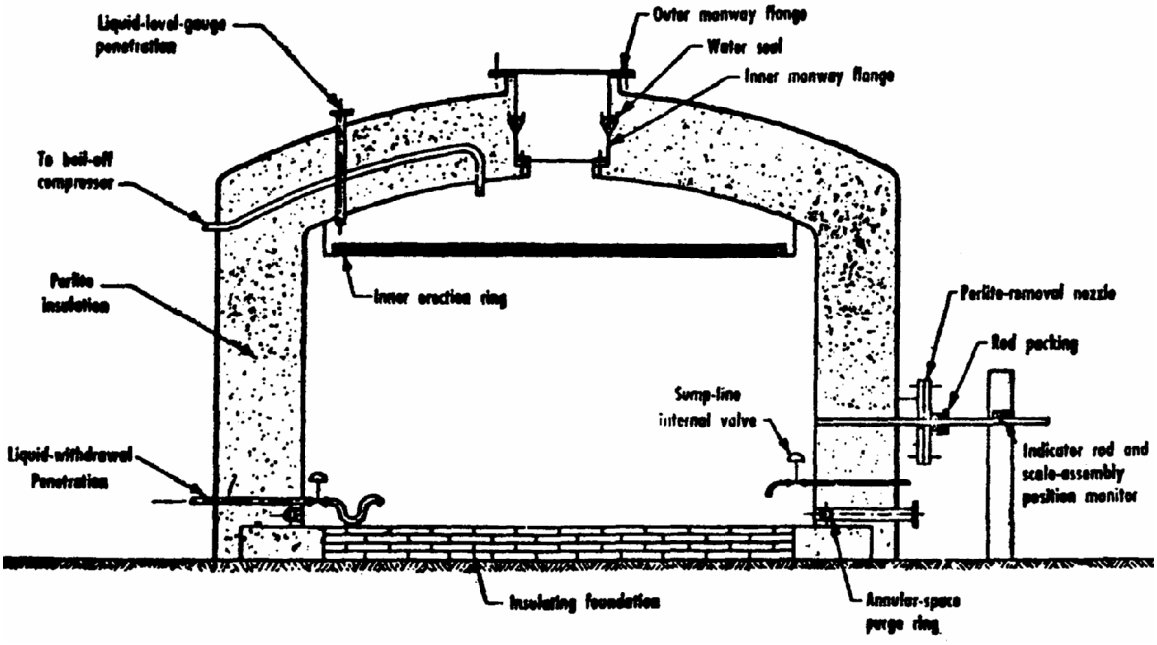
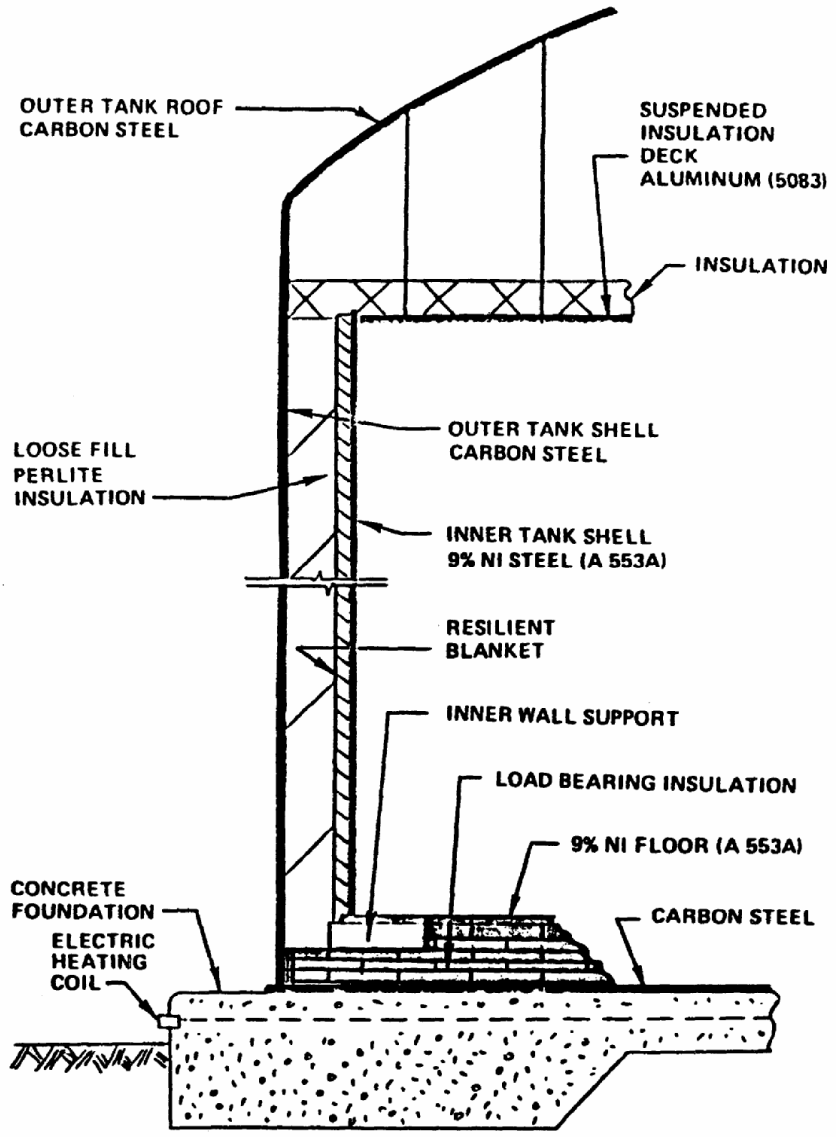
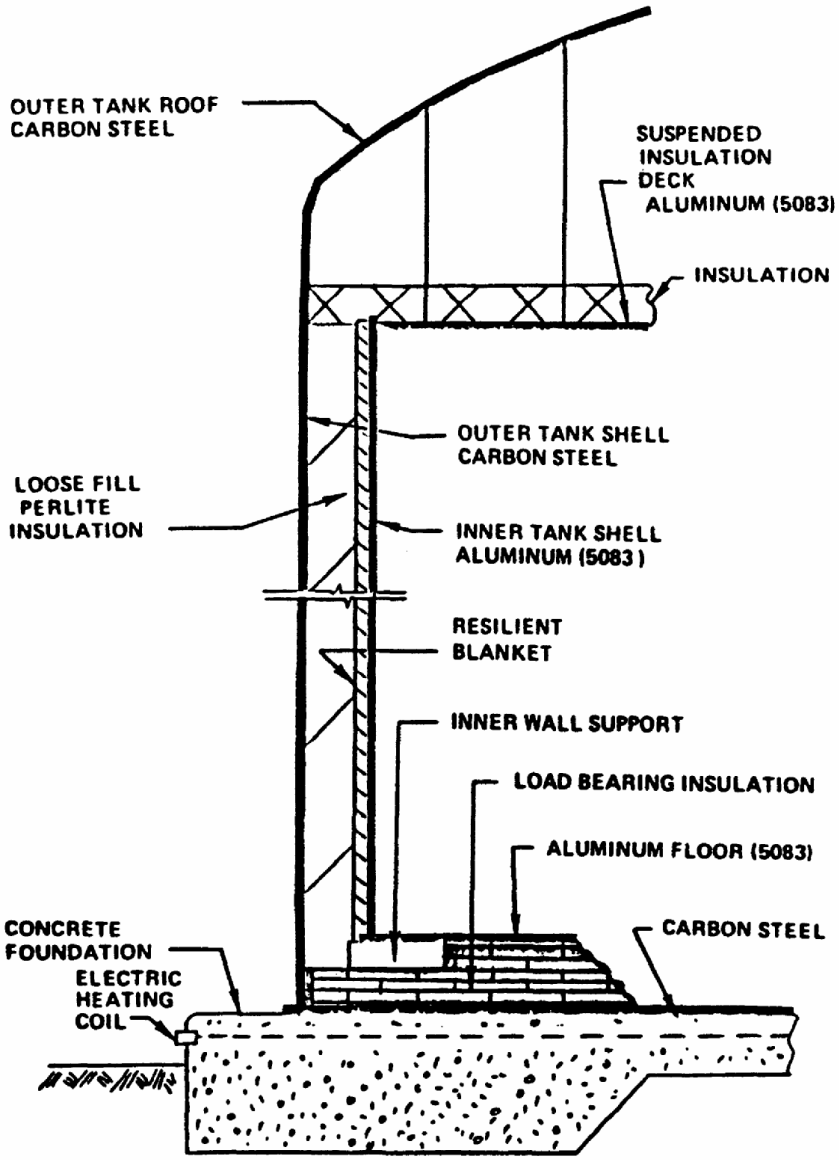


Figure 23.12: LNG Tank Cross Section



Double walled metallic tank with 9% nickel inner and carbon steel outer shells. It is the most common design utilized and is used widely by CBI, PDM and others. Over 70 tanks, 290m bbl and larger, have been constructed and are in operation throughout the world.

Figure 23.13: Alternate No. 1 LNG Storage Tank Design

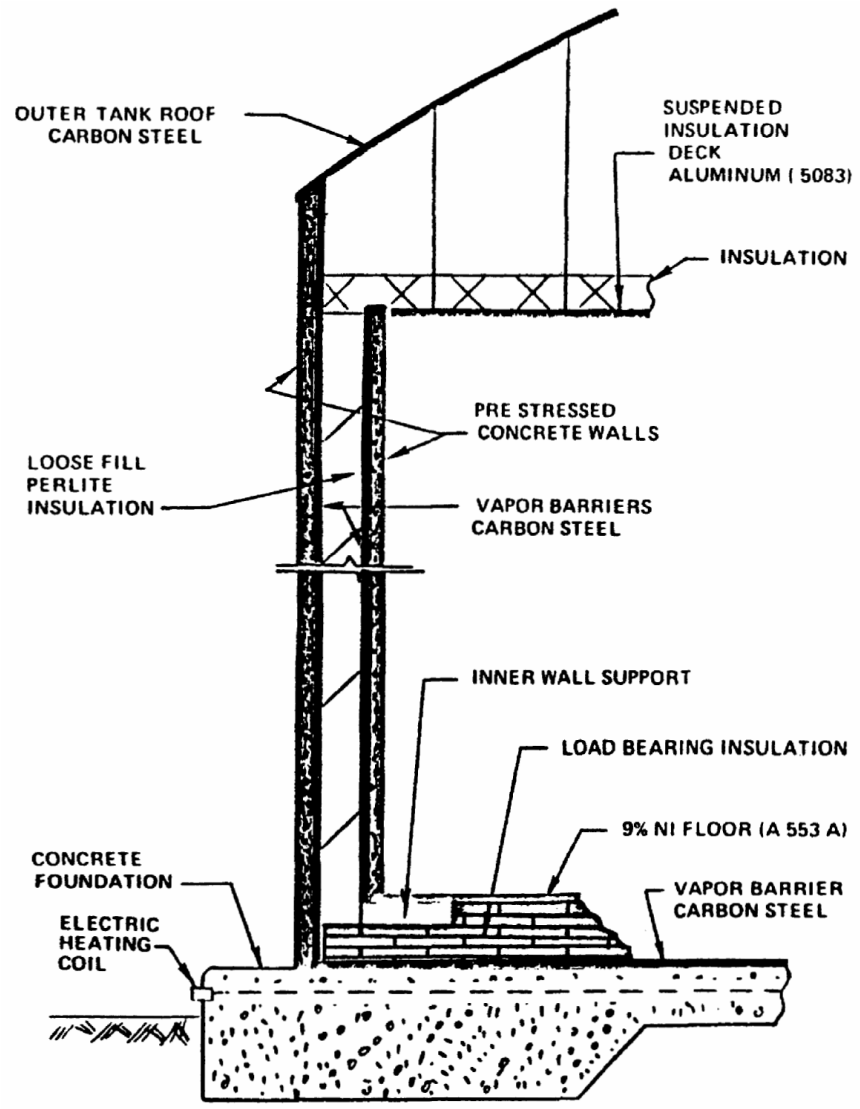


Double walled metallic tank with aluminum inner and carbon steel outer shells. Both CBI and PDM have designed and constructed tanks employing this concept.

Figure 23.14: Alternate No. 2 LNG Storage Tank Design

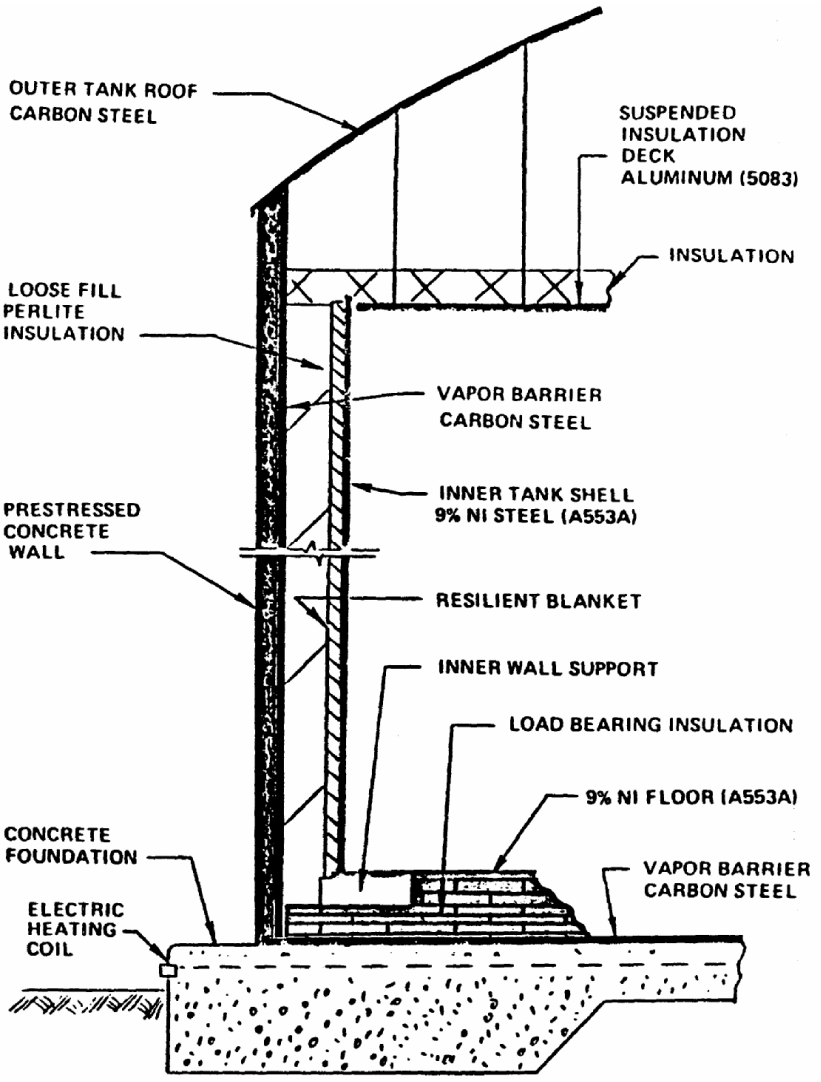
23.6.1.1 Prestressed Concrete Tanks

Very large tanks (800,000 BBLS size and greater) have been proposed. In these tanks an outer prestressed concrete shell with a vapour barrier on the inside is used along with an inner shell of low temperature resistant metal or another concrete inner shell. Figure 23.15, Figure 23.16, and Figure 23.17 show details of this design. Extra safety is claimed for this design because the outside wall can handle the full LNG load at the low temperature. Also it is claimed that a dyke is not needed. This has not been readily accepted.



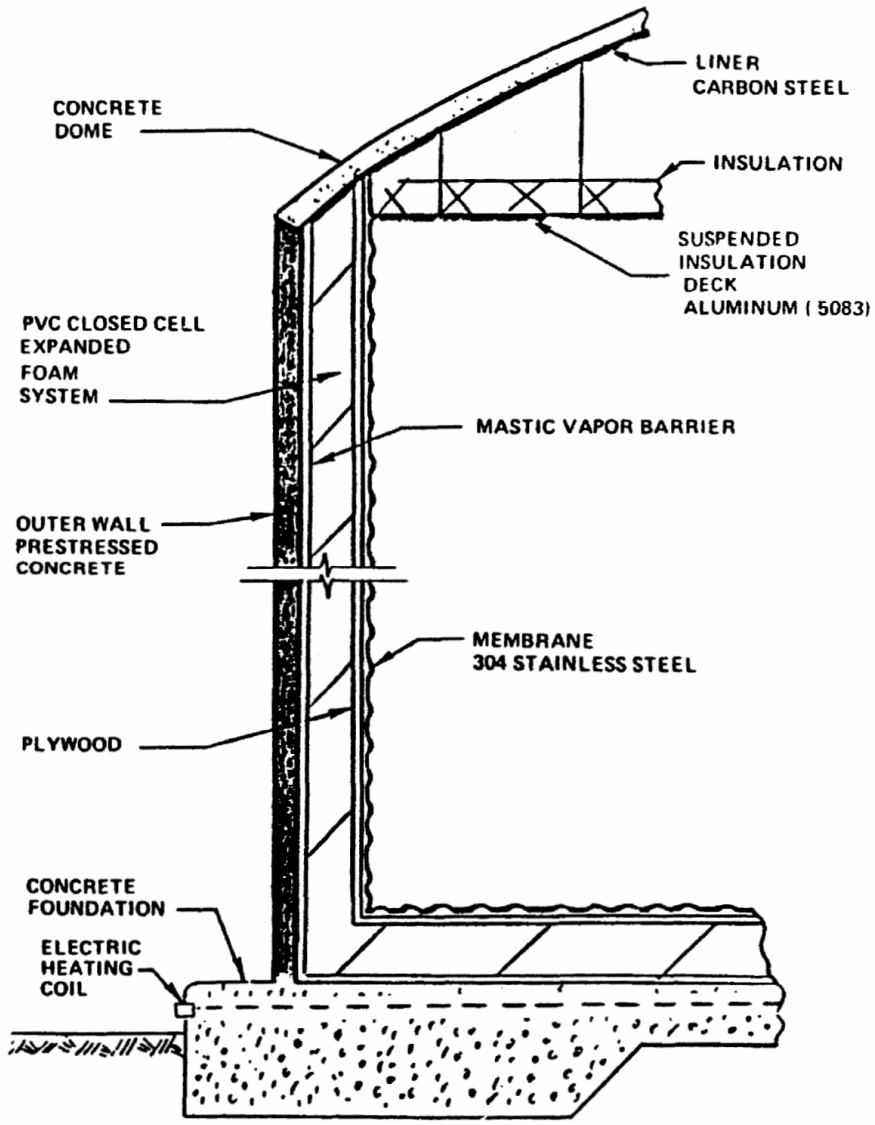
This concept presents a double prestressed concrete walled tank, both walls designed to contain the full tank of LNG.

Figure 23.15: Alternate No. 6 LNG Storage Tank Design



The concept provides an inner 9% nickel steel LNG containment system and an outer prestressed concrete shell, both designed to withstand the load of the full tank of LNG.

Figure 23.16: Alternate No. 8 LNG Storage Tank Design



This concept utilizes a prestressed concrete outer wall to act as a secondary, full integrity containment. This outer wall also provides protection against external impact sources. A waffle-design, stainless steel membrane liner is provided as the internal system in contact with the LNG product.

Figure 23.17: Alternate No. 11 LNG Storage Tank Design

Table 23.6 gives the approximate number of the types of LNG storage tanks that had been build up to, and including, 1979.

Table 23.6: Aboveground LNG Tank Alternatives

DESCRIPTION INNER/OUTER	SIZE MBBL	NO. INSTALLED	PROMINENT LOCATIONS	CONTRACTOR	OBSERVATIONS
9 Ni/CS	45-960	Approx. 100	Das Island	CBI/PDM/IHI	Since 1965
Al/CS	16-400	Approx. 70	Cove Point	CBI/PDM	Since 1963
9 Ni/CS/PC	348	1	Penna.	CBI	1972
9 Ni/9 Ni	——	0	Staten Is.	——	Proposed
9 Ni/9 Ni/RC	290	1	Astoria, NY	PDM	1974
PC/PC	500	1	Barcelona	Preload	U.C.
PC/Berm/PC	583-900	4	Phila/NY	Preload	1974/N.I.S.
9 Ni/PC	500	1	Fos-sur-Mer	Freyssinet	U.C.
Al/PC	——	0	——	——	Proposed
PC/CS	250	2	Barcelona	Preload	1969
SS Memb/PC	750	2	S. Nazaire	Freyssinet	U.C.
Invar Memb/PC	——	0	——	——	Proposed
Al Sphere	——	0	——	——	Proposed

23.7 Terminal Design

The selection and installation of the appropriate storage tanks is a major part of the design of terminals. There are several other items to consider. The transfer lines and pumps are very large to accommodate the loading of the large LNG carriers - 125,000 m^3 in size, such that they can be loaded in 12 hours (or unloaded). Also, the type of pumps used is quite important as they are generally submersible in the storage tanks. The return of vapour has to be considered closely. The joining of the loading or unloading lines to the gas sheds is also important. The actual design of the dock which must allow for proper berthing, preferably in both directions, also requires special engineering knowledge. The article by C. Durr and D.B. Crawford gives some data on terminal design.

An important consideration in terminal design and in storing LNG is the phenomenom of "Rollover" which is when the bottom portion of the liquid in a cryogenic storage tank is superheated because of a heat leak, then migrates to the surface rapidly and creates a large volume of vapour. The addition of LNG of different densities to a partially filled tank can cause the phenomena to occur. Proper mixing can prevent this problem.

23.8 LNG Marine Transportation

LNG is transported in very large ocean going carriers that are nominal 130,000 m^3 in size. Figure 23.18 shows a typical carrier. They are over 900 ft. long and have draught of 36 feet or greater. The design of such a vessel is a particular science and is primarily the province of naval architects and marine engineers (Good reference articles are by O.W.D. Thomas and A.H. Schwendtner and

B.D. Frondevillei). However the type of storage system is of considerable interest to the gas engineer. There are two types of cargo systems:

- Free standing or self supporting tanks that have sufficient strength to stand by themselves and to support the weight and dynamic forces of the cargo.
- Membranes or integral tanks in which a thin metal barrier supported by insulation contains the liquid. The membrane tank insulation transmits the weight and dynamic forces of the cargo to the inner hull structure of the hull itself.

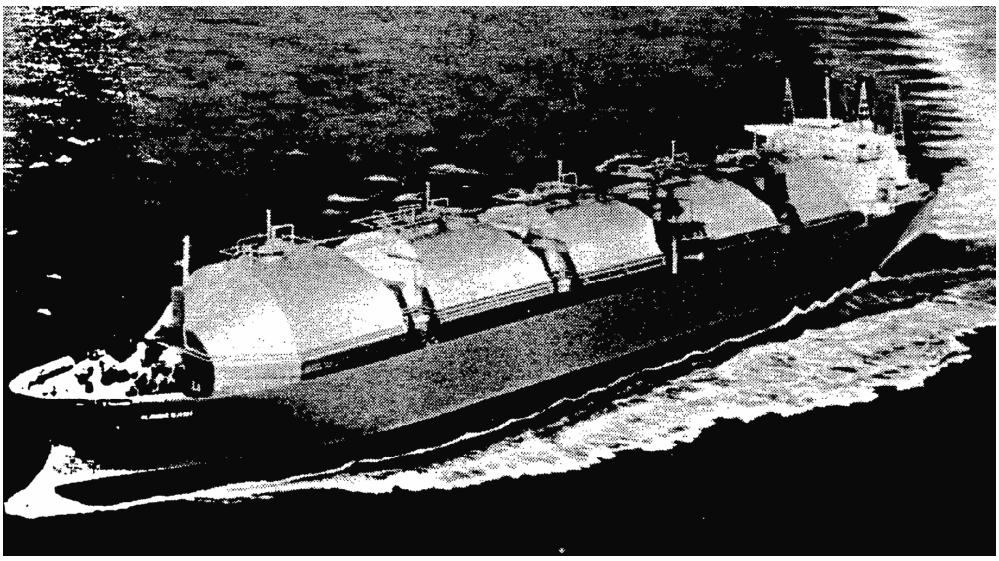


Figure 23.18: LNG Ocean Going Carrier

23.8.1 Freestanding Cargo Tanks

There are two basic types of freestanding tanks, one prismatic, the other spherical in shape.

The prismatic tank, designed by Conch International Methane, Ltd., is basically a flat-walled conventional tank structure stiffened by vertical members in turn supported by deep horizontal webs and diagonal struts. It is built of aluminum. The design has been optimized from the standpoints of weight and ease of construction. It has been greatly refined by the use of finite element analysis techniques.

The insulation system consists mainly of reinforced polyurethane foam and layers of fibre glass. The characteristics of the polyurethane produce a liquid-tight secondary barrier except at the top of the tank, where it is not required.

Spherical tanks, developed by Kvaerner-Moss, Sener, and Chicago Bridge and Iron, and distinguished by "leak-before-failure" or "fail-safe" design concepts. They are constructed of either nickel-steel or aluminum. The shape of the tanks is especially suitable for evaluation by the techniques of fracture mechanics and stress analysis, which lead to a confidence level so high that the structural integrity of a spherical tank can be virtually guaranteed.

Basically, the insulation system consists of polyurethane or polystyrene foam contiguous to the entire outer surface of the sphere. Figure 23.19 shows a spherical tank design.

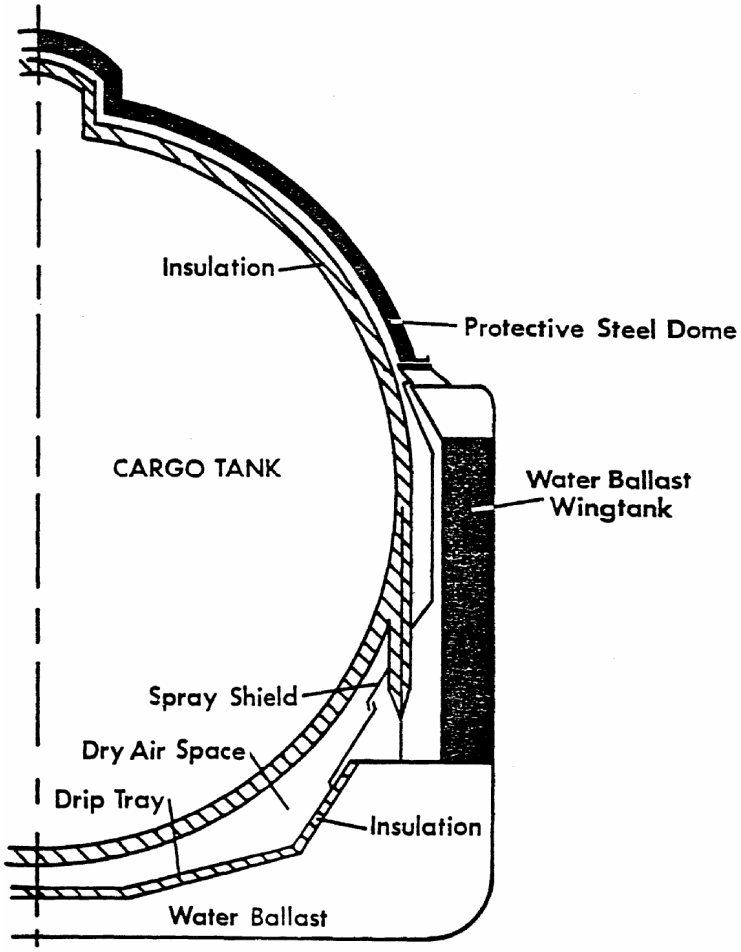


Figure 23.19: Spherical Tank Design

23.8.2 Membrane Tanks

There are two leading membrane tank designs, one developed by Gas-Transport, the other by Technigaz.

In the Gas-Transport design of two layers of insulation and two membrane barriers of a metal alloy, one layer of insulation covers the interior (cargo) side of the inner hull of the ship. The insulation is covered, in turn, by a metal alloy membrane secondary barrier. The barrier is covered by a second layer of insulation, which is, in turn, covered by the membrane primary barrier.

Both layers of insulation are essentially identical, consisting of plywood boxes filled with perlite. The two metal barriers are also essentially identical. The metal is Invar, a nickel-iron alloy that expands or contracts very little when exposed to the temperature range experienced in LNG transport. Each barrier, which is only 0.02 inches thick, is applied in 16 inch wide strips and is joined by means of automatic welding devices.

In the Technigaz design, the tank structure consists of a stainless steel primary membrane barrier and a sealed plywood secondary membrane barrier. The spaces between the secondary barrier and the inner hull of the ship are insulated with laminated balsa panels. Balsa pads are bonded to the face of the plywood, and to these is fastened the primary barrier. The Technigaz design is distinguished by the waffled configuration of the primary membrane.

Figure 23.20 shows a membrane tank design.

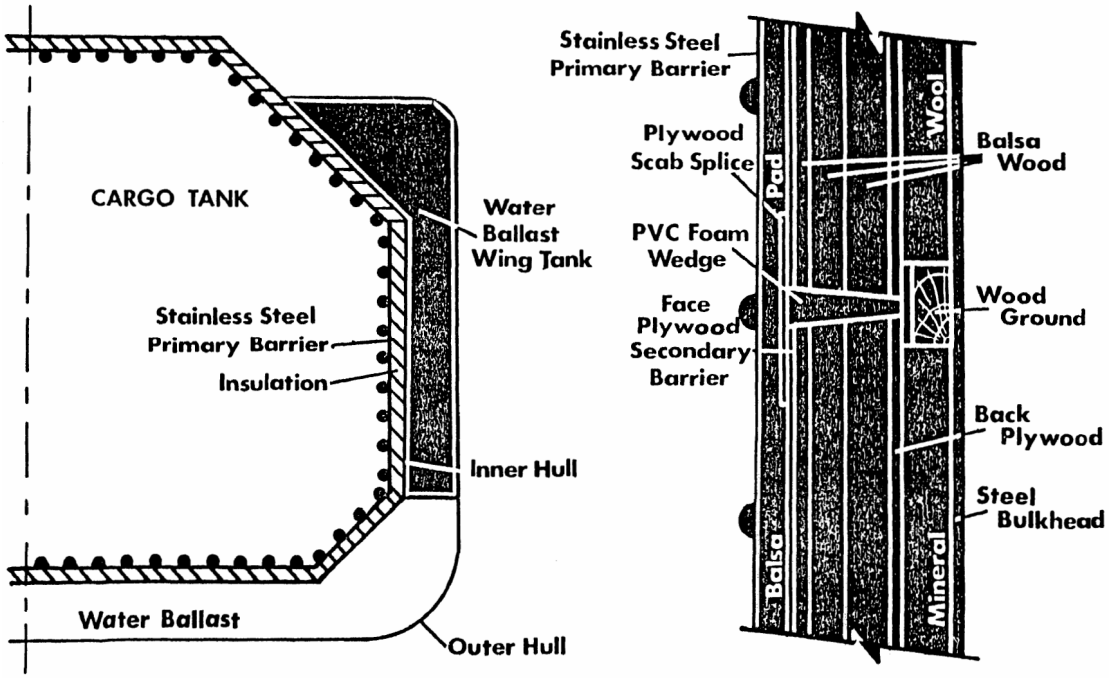


Figure 23.20: Membrane Tank Design

23.8.3 Fleet Size

The LNG carrier of today represents the most expensive cargo vessels ever constructed for the maritime trade. In addition to the high initial capital costs for construction, the LNG carrier operational expenses are also a significant input to the economic evaluation of the viability of an LNG project. Every effort must be made by the LNG project sponsors to ensure an accurate evaluation of the LNG carriers capabilities and to optimize its use in the overall performance of the liquefaction, transportation and regasification facilities of the total project.

To evaluate the interrelated shipping and terminal requirements for LNG projects, other LNG transportation project sponsors have developed an LNG shipping trade computer model program. This simulation model permits simultaneous consideration of the liquefaction plant production rate and storage facilities, loading terminal berthing and transfer requirements, fleet operations and transportation capacity, unloading terminal facilities, and receiving terminal storage capacity and vaporization rate.

Complex problems of evaluating the interfacing requirements of these various facilities develop when a variable flow process, the marine transportation element, is inserted between two essentially constant flow processes, the liquefaction and regasification plants.

Definition of the variable flow process, that is the operation of the LNG tanker fleet, can be made by evaluating each step in the round trip circuit of each ship between the loading and unloading terminals plus allowances for periodic and unscheduled maintenance and repairs for the vessels. This is accomplished by a series of inputs to the computer program.

A simplified schematic of the simulation process is shown in Figure 23.21.

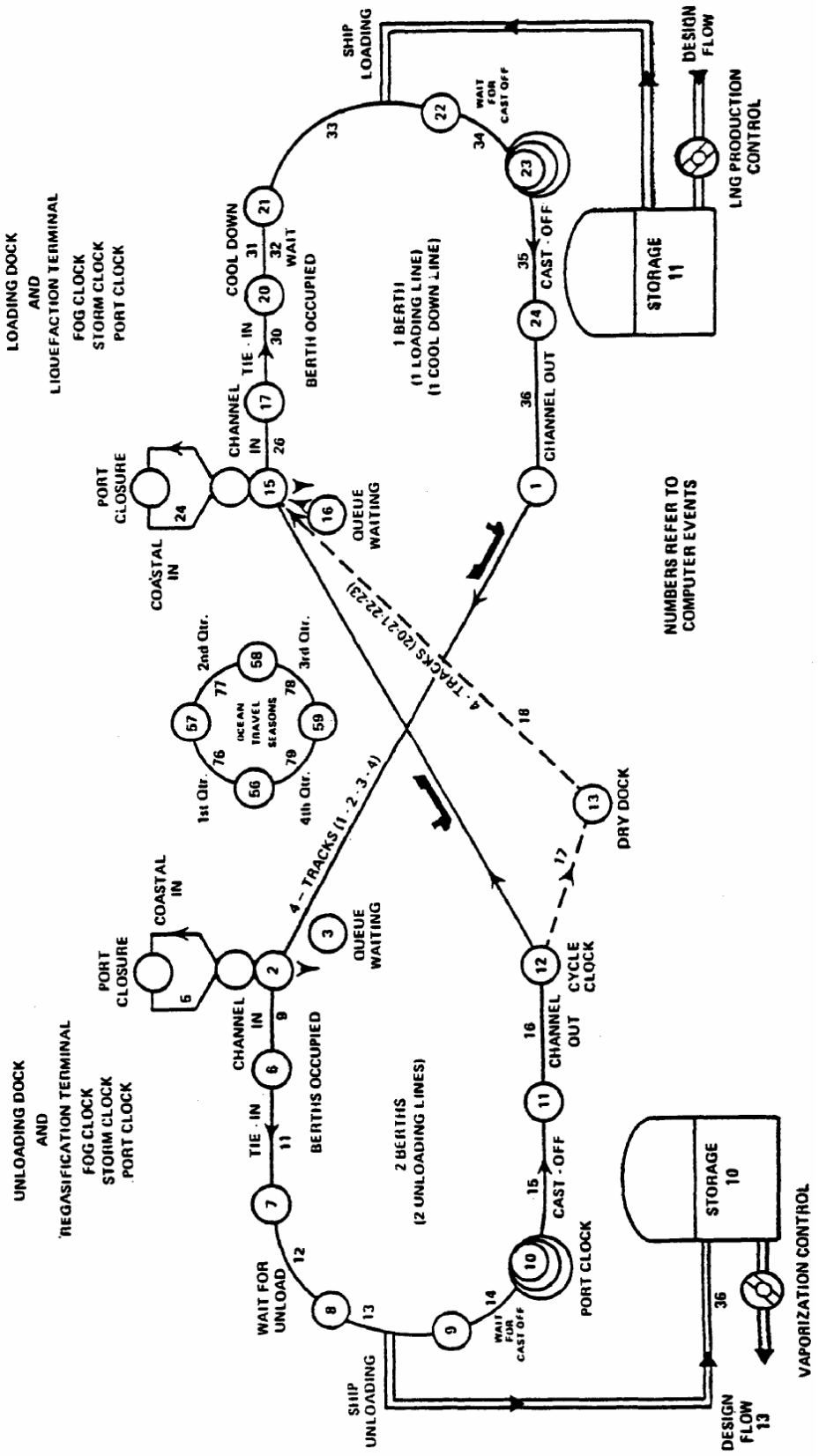


Figure 23.21: Two Port Simulator Schematic - Shipping Model

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24.0 Gas Handling and Processing Summary

- An outline how gas is generally handled depending on the volume being produced and its composition from the wellhead to delivery as a specific product to a transportation system.

24.1 Introduction

The next few pages presents a summary of how gas could be processed to make pipeline quality gas based primarily on technical considerations. Varying qualities of gas from lean sweet gas to rich sour gas are considered. Two sizes of facilities are considered; small, below 1.5 x 10^6 m^3/day (53 MMSCFD) and large, greater than 1.5 x 10^6 m^3/day (53 MMSCFD). Generally, it is expected that the gas will be the non-associated type at reasonable pressures. Low pressures that require compression or solution gas facilities will require a close look at the suggested methods. Similarly, a specific review of the economics of the actual project is needed before the final gas production gathering and process scheme, particularly if the production of propane, butane, etc. instead of NGL's is selected. Thus the attached tables should only be used as a guide.

Table 24.1: Raw Gas Quality Description

Hydrocarbon Content	
Lean	Mostly C_1, C_3^+ 2.0% or lower (HC basis), C_5^+ 0.1% or lower.
Medium	C_3^+ up to 5.0% (HC basis), C_6^+ 0.3% or lower, needs HC dewpoint control unit.
Rich	Anything above previous listed criteria.
Acid Gas Content	
Sweet	Meets pipeline gas specs, ie. H_2S < 16 ppm (or 4 ppm), CO_2 less than 2%.
High CO_2 Trace H_2S	Greater than 2% CO_2, H_2S up to 200 ppm.
Sour (both H_2S & CO_2)	Above previous listed criteria.

Table 24.2: Small Gas Facilities (to 1.5 x 10^6 m^3/day - 53 MMSCFD)

Gas Quality Description					
Hydrocarbon	**Acid Gas**	**Wellhead Treating Likely Options**	**Gathering System Comments**	**Processing Options and Discussion**	**Products Recovered**
Lean	Sweet	Mostly dehydrators. May be glycol or methanol hydrate inhibition to get to plant. Heaters could be used for short lines.	Nothing special - follow regular gas code. High % value of yield where applicable.	Wellhead dehydration may enough or central dehydration if have hydrate inhibition at wellhead.	Gas only.
Lean	High CO_2 Trace H_2S	Mostly dehydrators or glycol or methanol inhibition. Heaters could be used for short lines.	Follow regular gas code. High % value of yield where applicable. Consider inhibition.	Consider CO_2 removal process, ie. hot potassium carbonate (No H_2S) or sulphinol, selexol, etc.; also look at chemical solvents. H_2S scavenging process for trace H_2S. Glycol dehydration for water dewpoint.	Gas only.

Gas Quality Description					
Hydrocarbon	**Acid Gas**	**Wellhead Treating Likely Options**	**Gathering System Comments**	**Processing Options and Discussion**	**Products Recovered**
Lean	Sour (both H_2S & CO_2)	Primarily line heaters. Could use dehydrators if significantly cheaper and H_2S not too high.	Need special low carbon soft steel, max. yield lower, inhibit lines starting at wellhead.	Look at physical solvents, sulphinol, selexol and chemical solvents. Claus unit to 5 t/d, consider LoCat, Stretford below this. Processes glycol dehydration for water dewpoint.	Gas and sulphur.
Medium	Sweet	Primarily dehydration. Look at bypassing liquids around dehydrator. Look at heaters if have short lines.	Nothing special as far as codes. However probably have two phase lines - pigging facilities probably not needed, standard inlet separator.	Refrigeration dewpoint control and condensate stabilization.	Gas and condensate (C_5^+).
Medium	High CO_2 Trace H_2S	Probably still dehydration. Look at bypassing liquids around dehydrator. Look at heaters if have short lines.	Nothing special as far as codes, have two-phase lines - pigging facility probably not needed. Standard inlet separator. Consider inhibition. Both phases.	Consider CO_2 removal processes. Hot potassium carbonate for CO_2 only or sulphinol, selexol, etc. Also look at chemical solvents. H_2S traces - scavenging processes then refrigeration dewpoint control.	Gas and condensate (C_5^+).
Medium	Sour (both H_2S & CO_2)	Line heaters.	Need special low carbon soft steel. Have two phase lines. Inhibit with both gas phase and liquid phase inhibitors. Standard inlet separator.	Look into both physical and chemical solvents i.e. sulphinol, DEA, etc. Do study on MDEA if H_2S selective needed. Refrigeraton dewpoint control. Claus unit to 5 tons, consider LoCat, Stretford below that.	Gas and condensate (C_5^+), Sulphur.
Rich	Sweet	Line heaters.	Nothing special, have two phase flow, put in pigging facilities. Size separator for large slugs.	Refrigeration dewpoint control or small expander type unit depending on size. Have a deethanizer with this.	Gas NGL's which include C_5^+
Rich	High CO_2 Trace H_2S	Line heaters.	Nothing special - have two phase flow, put in pigging facilities. Size separator for large slugs. Consider inhibition both phases.	Probably consider both physical and chemical processes - check by circulation rate. Refrigeration dewpoint control or small expander depending on size. Have deethanizer with this.	Gas NGL's which include C_5^+
Rich	Sour (both H_2S & CO_2)	Line heaters.	Need special low carbon soft steel, max. yield lower, inhibit lines - both liquid and gas phases. Have two phase flow, put in pigging facilities. Size separator for large slugs. Inhibit both phases.	Consider primarily and chemical processes check by circulation rates. Refrigeration dewpoint control or small expander depending on size. Have a deethanizer with this. Claus unit to 5 tons, consider LoCat, Stretford below that.	Gas, Sulphur and NGL's which include C_5^+

Table 24.3: Large Gas Facilities (greater then 1.5 x 10^6 m^3/day - 53 MMSCFD)

Gas Quality Description					
Hydrocarbon	**Acid Gas**	**Wellhead Treating Likely Options**	**Gathering System Comments**	**Processing Options and Discussion**	**Products Recovered**
Lean	Sweet	Mostly glycol dehydrators. May be	Nothing special - follow regular gas code. High %	Plant dehydration via glycol.	Gas only.

Gas Quality Description					
Hydrocarbon	**Acid Gas**	**Wellhead Treating Likely Options**	**Gathering System Comments**	**Processing Options and Discussion**	**Products Recovered**
		glycol or methanol hydrate inhibition to get to plant. Heaters could be used for short lines.	value of yield where applicable.		
Lean	High CO_2 Trace H_2S	Mostly glycol dehydration or glycol or methanol inhibition. Heaters could be used for short lines.	Follow regular gas code. High % value of yield where applicable. Consider inhibition.	Consider CO_2 removal process, ie. sulphinol, selexol, Hot Pot if no CO_2. Also look at chemical solvents. H_2S scavenging if required.	Gas only.
Lean	Sour (both H_2S & CO_2)	Primarily line heaters. Could use dehydrators if significantly cheaper and H_2S not too high.	Need special low carbon soft steel, max. yield lower, inhibit lines starting at wellhead.	Look at physical solvents, sulphinol, selexol and chemical solvents. Possibly MDEA if high CO_2 low H_2S. Have a Claw plant to 5 tons. Consider LoCat, Stretford below that. Glycol drhydration for water dewpoint.	Gas and sulphur.
Medium	Sweet	Mostly glycol dehydration. Look at bypassing liquids around dehydrator. Look at heaters for short lines.	Nothing special as far as codes. However probably have two phase lines - pigging facilities probably not needed, standard inlet separator.	Refrigeration dewpoint control, condensate stabilizer or deethanizer depending on NGL volume.	Gas Condensate (C_5^+), NGL's possibly.
Medium	High CO_2 Trace H_2S	Probably still glycol dehydration. Look at bypassing liquids around dehydrator. Heaters could be used for short lines.	Nothing special as far as codes, have two-phase lines - pigging facility probably not needed. Standard inlet separator.	Look at both, physical and chemical solvents sulphinol, selexol, DEA. Hot Pot if CO_2 only. Refrigeration dewpoint control. Condensate stabilizer or deethanizer depending on NGL volume.	Gas Condensate (C_5^+), NGL's possibly.
Medium	Sour (both H_2S & CO2)	Line heaters.	Need special low carbon soft steel. Have two phase lines. Inhibit lines with both gas phase and liquid phase inhibitors. Standard inlet separator.	Look into both physical and chemical solvents ie. sulphinol, DEA, etc. MDEA if H_2S selectivity needed. Refrigeration dewpoint control. Have Claus unit to 5 tons; consider LoCat or Stretford below that.	Gas Condensate (C_5^+), NGL's possibly, Sulphur.
Rich	Sweet	Line heaters.	Nothing special, have two phase flow, put in pigging facilities. Size separator for large slugs.	Expander for NGL recovery. Fractionation for NGL products.	Gas, NGL's or spec, C_2 - review market, C_3, C_4, C_5^+
Rich	High CO_2, Trace H_2S	Line heaters.	Nothing special - have two phase flow, put in pigging facilities. Size separator for large slugs. Consider inhibition for both phases.	Look at physical solvents, ie. sulphinol, selexol. Hot Pot C02 only, and chemical solvents. Select by circulation rates. Expander for NGL recovery. Fractionation for NGL products.	Gas, Sulphur, NGL's or spec, C_2 - review market, C_3, C_4, C_5^+
Rich	Sour (both H_2S & CO_2)	Line heaters.	Need special low carbon soft steel, max. yield lower, - inhibit lines both liquid and gas phases. Have two phase flow, put in pigging facilities. Size separator for large slugs.	Consider primarily and chemical solvents. Check by circulation rate. Expander for NGL recovery. Fractionation for NGL products. Claus unit to 5 tons; consider LoCat or	Gas, Sulphur, NGL's or spec, C_2 - review market, C_3, C_4, C_5^+

Gas Quality Description					
Hydrocarbon	**Acid Gas**	**Wellhead Treating Likely Options**	**Gathering System Comments**	**Processing Options and Discussion**	**Products Recovered**
				Stretford below that.	

Appendix 2.0 Natural Gas Processing Principles and Technology - Part II

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Reference of Key Equations

$$U_2 + \frac{v_2^2}{2g_C} + Z_2 + P_2V_2 = U_1 + \frac{v_1^2}{2g_C} + Z_1 + P_1V_1 + Q + W \quad \text{.....10-3}$$

$$\Delta U + \frac{\Delta\left(v_2^2 - v_1^2\right)}{2g_C} + \Delta Z + \Delta PV = Q + W \quad \text{.....10-4}$$

$$W = \int_P^{P_2} V dP \quad \text{.....10-4}$$

$$HP = 0.004363 P_1 V_1 \ln\left(\frac{P_2}{P_1}\right) \quad \text{10-4}$$

$$Work = \Delta H - T\Delta S \quad \text{10-4}$$

$$HP = 0.0432(\Delta H - T\Delta S) \quad \text{10-5}$$

$$HP = 0.004363 P_1 V_1 \left[\left(\frac{K}{K-1}\right)\left(\frac{P_2}{P_1}\right)^{K-1/K} - 1\right] \quad \text{10-5}$$

$$Work = \Delta H \quad \text{10-5}$$

$$HP = 0.0432 \Delta H \quad \text{10-5}$$

$$BHP = BHP \Big/ MMSCFD\left(\frac{P_L}{14.4}\right)\left(\frac{T_S}{T_L}\right)(Z_{ave}) \qquad (Flow\ in\ MMSCFD) \quad \text{10-6}$$

$$W = \frac{W}{m^3/hr}\left(\frac{P_L}{101.325}\right)\left(\frac{T_S}{T_L}\right)(Z_{ave}) \qquad (Flow\ in\ m^3/hr) \quad \text{10-6}$$

$$MC_P - MC_V = 1.99 \quad \text{10-7}$$

$$K = \frac{MC_P}{MC_P - 1.99} \quad \text{10-7}$$

$$K = \frac{MC_P}{MC_P - 8.3145} \quad \text{10-7}$$

$$P_D = \frac{A_{HE}(S)(RPM)}{1728} \quad \text{10-9}$$

$$P_D = \frac{A_{HE}(S)(RPM)(60)}{4\ x\ 10^9} \quad \text{10-9}$$

$$P_{DA} = \frac{(S)(RPM)(2)}{1728}\left[A_{HE} - \frac{A_R}{2}\right] = 4.55\ x\ 10^{-4}(S)(RPM)(2D^2 - d^2) \quad \text{10-9}$$

$$P_{DA} = \frac{(S)(RPM)\,2\left(A_{HE} - A_R/2\right)}{4\ x\ 10^9} \quad \text{10-9}$$

$$Capacity = P_D(E_V)\left(\frac{P_1}{14.4}\right)(1440)\left(\frac{520}{T_S}\right)\left(\frac{1}{Z}\right) \quad \text{10-10}$$

$$Capacity = P_D E_V P_1(100) \quad \text{10-10}$$

$$m^3/hr = \frac{P_D E_V P_1}{(100)(101.325) Z_{ave}} \quad \text{10-10}$$

$$E_V = 96 - R - \%Cl\left[\left(\frac{Z_S}{Z_d}\right)(R)^{1/K} - 1\right] \quad \text{10-10}$$

$$Rod\ Load\ in\ compression = A_{HE} P_2 - A_{CE} P_1 \quad \text{10-11}$$

$$BHP/MMSCFD = 240.5\left[(R)^{\frac{K-1}{K}} - 1\right] \quad \text{10-14}$$

$$BHP = \frac{1545(Z_a) T_1 W}{33000\, E(MW)}\left(\frac{K}{K-1}\right)\left[(R)^{\frac{K-1}{K}} - 1\right] \quad \text{10-14}$$

$$Power = \frac{1000W\left(\frac{8.314}{M}\right)\left(\frac{K}{K-1}\right)(Z_{ave} T_1)\left[\left(\frac{P_2}{P_1}\right)^{\frac{K-1}{K}} - 1\right]}{3600000(E)} \quad \text{10-14}$$

$$H_{adiabatic} = ZRT\left(\frac{K}{K-1}\right)\left[\left(\frac{P_2}{P_1}\right)^{\frac{K-1}{K}} - 1\right] \quad \text{10-15}$$

$$H_{adiabatic} = \psi \frac{u^2}{g} \quad \text{10-15}$$

$$Q\ \alpha\ N;\ i.e.\ \frac{Q_{110}}{N_{110}} = \frac{Q_{100}}{N_{100}} = \frac{Q_{90}}{N_{90}} \quad \text{10-16}$$

$$H\ \alpha\ N^2;\ i.e.\ \frac{H_{110}}{(N_{110})^2} = \frac{H_{100}}{(N_{100})^2} = \frac{H_{90}}{(N_{90})^2} \quad \text{10-16}$$

$$Bhp\ \alpha\ N^3;\ i.e.\ \frac{Bhp_{110}}{(N_{110})^3} = \frac{Bhp_{100}}{(N_{100})^3} = \frac{Bhp_{90}}{(N_{90})^3} \quad \text{10-16}$$

$$Power = (H_1 - H_2) E \quad \text{10-18}$$

$$\beta = \frac{Hb - He}{Hc - Hb} \quad \text{10-22}$$

$$HP\ per\ ton\ of\ refrigeration = \frac{4.713}{\beta} \quad \text{10-23}$$

$$\frac{Theoretical\ HP}{0.80} \quad \text{10-23}$$

$$\%\ Risk = \frac{(Original\ Net\ Cost + Cost\ of\ Way\ Out - Net\ Cost\ of\ Safe\ Plan)}{Net\ Cost\ of\ Safe\ Plan - Original\ Net\ Cost} \times 100\% \quad \text{11-2}$$

$$New\ Cost = Cost\ of\ Existing\ Unit\left(\frac{size\ of\ new\ unit}{size\ of\ old\ unit}\right)^{0.6} \quad \text{11-19}$$

$$Q = C(P_f^2 - P_s^2)^n \quad \text{12-1}$$

$$Q = C'(Psiw^2 - Pw^2)^{n'} \quad \text{12-7}$$

$$Q\max = qt\left[\frac{P_{FO}^2 - f^2 P_{FO}^2}{P_F^2 - P_S^2}\right]^n \left[\frac{P_F}{P_{FO}}\right] \quad \text{12-12}$$

$$q = UA\frac{(T_1 - t) - (T_2 - t)}{\ln\left(\frac{T_1 - t}{T_2 - t}\right)} = UA(\Delta t_{lm}) \quad \text{13-4}$$

$$q = W(H_1 - H_2) \quad \text{13-5}$$

$$A = \frac{W(H_1 - H_2)}{U(\Delta t_{lm})} \quad \text{13-5}$$

$$d = \frac{KW}{100\,M - MW} \quad \text{13-10}$$

$$Q = \frac{5634 d^{2.5}(T_b + 460)}{P_b(STlZ)^{1/2}} \frac{1}{\sqrt{f}} \left(P_1^2 - P_2^2\right)^{1/2} \quad \text{14-5}$$

$$Q = 0.1034 \frac{d^{2.5}}{(GTLZ)^{0.5}} \frac{1}{\sqrt{f}} \left(P_1^2 - P_2^2\right)^{0.5} \quad \text{14-5}$$

$$Q_D = 433.5 \frac{T_b}{P_b}\left[\left(\frac{P_1^2 - P_2^2}{STlZ}\right) d^{16/3}\right]^{1/2} \quad \text{14-6}$$

$$Q_D = 737\left(\frac{T_b}{P_b}\right)^{1.02} E\left[\frac{P_1^2 - P_2^2}{G^{0.961} LTZ}\right]^{0.51} d^{2.53} \quad \text{14-7}$$

$$Q_D = 0.0037 \frac{T_b}{P_B}\left[\left(\frac{P_1^2 - P_2^2}{GTLZ}\right) d^{16/3}\right]^{1/2} \quad \text{14-7}$$

$$Q_D = 0.010 E\left(\frac{T_b}{P_b}\right)^{1.02} \left(\frac{P_1^2 - P_2^2}{G^{0.961} LTZ}\right)^{0.510} d^{2.53} \quad \text{14-7}$$

$$\Delta P_{TOTAL} = \Delta P_{TP} + \Delta P_h \quad \text{14-8}$$

$$\Delta P_h = \sum \frac{hF_e \rho_C}{144} \quad \text{14-8}$$

$$\Delta P_h = \sum 0.009807\, hF_e \rho \quad \text{14-9}$$

$$P = \frac{2StFET}{d} \quad \text{14-18}$$

$$P = \frac{2St}{D} \times 10^3\, FET \quad \text{14-20}$$

$$CE = C + \frac{Mn}{6} + \frac{Cr + Mo + Va}{5} + \frac{Ni + Cu}{15} \quad \text{14-21}$$

$$V_g = \sqrt{\frac{2g\, M_p \left(\rho_l - \rho_g\right)}{\rho_l \rho_g A_p C'}} = \sqrt{\frac{4gD_p \left(\rho_l - \rho_g\right)}{3\rho_g C'}} \quad \text{15-7}$$

$$\mathrm{Re} = \frac{1488\, D_p V_t \rho_g}{\mu} \quad \text{15-7}$$

$$C'(\mathrm{Re})^2 = \frac{(0.95)(10^6)\rho_g D_p^3 \left(\rho_l - \rho_g\right)}{\mu^2} \quad \text{15-7}$$

$$V_t = 1.74 \frac{gD_p \left(\rho_l - \rho_g\right)}{\rho_g} \quad \text{15-8}$$

$$D_p = K_{CR} \left[\frac{\mu^2}{g\rho_g \left(\rho_l - \rho_g\right)} \right]^{1/3} \quad \text{15-8}$$

$$V_t = \frac{1488 g D_p^2 \left(\rho_l - \rho_g\right)}{18\mu} \quad \text{15-9}$$

$$u = K \sqrt{\frac{\rho_L - \rho_g}{\rho_g}} \quad \text{15-11}$$

$$\frac{Q}{u} = A \quad \text{15-12}$$

$$d = \frac{4000W}{100M - MW} \quad \text{17-3}$$

$$d = \frac{2222W}{100M - MW} \quad \text{17-3}$$

$$\alpha = \frac{K_{\ell k}}{K_{hk}} \quad \text{18-1}$$

$$G = C\sqrt{\rho v(\rho L - \rho v)} \quad \text{18-2}$$

$$(\alpha_{LK})^{Sm} = \frac{(X_{LKD})(X_{HKW})}{(X_{LKW})(X_{HKD})} \quad \text{18-3}$$

$$\frac{\alpha_1 X_{1F}}{\alpha_1 - \theta} + \frac{\alpha_2 X_{2F}}{\alpha_2 - \theta} + \ldots\ldots\ldots = 1 - q \quad \text{18-3}$$

$$\frac{\alpha_1 X_{1D}}{\alpha_1 - \theta} + \frac{\alpha_2 X_{2D}}{\alpha_2 - \theta} + \ldots\ldots\ldots = R_{\min} + 1 \quad \text{18-3}$$

$$q = \frac{H_V - H_F}{H_V - H_L} \quad \text{18-3}$$

$$X = \frac{Q.10^6}{\pi U \sigma_y \sigma_z} e^{-\left(\frac{H^2}{2\sigma_z^2}\right)} \quad \text{22-10}$$

$$H = h_p + h_r - t - 3/4T \quad \text{22-12}$$

$$hr = \frac{C X^{\frac{2}{3}} F^{\frac{1}{3}}}{U} \quad \text{22-12}$$

$$F = \frac{g}{\pi}\frac{T_S - T_a}{T_a} Q_T \quad \text{22-13}$$

$$F = \frac{g}{\pi}\frac{Q_H}{C_p \rho T_a} \quad \text{22-14}$$

$$1500 = \frac{960W\sqrt{M}}{4\pi[H(H + 120D)]} \quad \text{22-15}$$

$$W_2 = W_1 + T_2 \Delta S + \Delta W_3 \quad \text{23-22}$$

$$W_1 = Q\frac{(T_2 - T_1)}{T_1} \quad \text{23-22}$$