

**ENGINEERING STANDARD**

**FOR**

**PROCESS DESIGN OF GAS TREATING UNITS**

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## 0. INTRODUCTION

"Design of Major, Non-Patented Refining Processes" are broad and contain subjects of paramount importance. Therefore a group of IPS Standards are prepared to cover the subject. The Process Engineering Standards of this group includes the following Standards:

<b><u>STANDARD CODE</u></b>	<b><u>STANDARD TITLE</u></b>
IPS-E-PR-491	"Process Requirements of Refinery Non-Licensed Units"
IPS-E-PR-500	"Process Design of LPG Recovery & Splitter Units"
IPS-E-RR-551	"Process Design of Gas Treating Units"

This Engineering Standard Specification covers:

### **"PROCESS DESIGN OF GAS TREATING UNITS"**

## 1. SCOPE

This Engineering Standard Specification is intended to cover the minimum process requirements, criteria and features for accomplishment of process design of gas treatment Units.

The primary operation of gas treatment process falls into gas sweetening and dehydration by absorption, adsorption and chemical reactions, so the general scope should be covered in two following parts, from which in this Standard only the part I has been discussed:

**Part I:** "Process Design of Gas Sweetening Units".

**Part II:** "Process Design of Gas Dehydration Unit".

The basic principles for process design of main equipment, piping and instrumentation together with guidelines on present developments and process selection are considered as the main objectives.

However, the nature and complexity of the problems, does not allow, that this Volume at this stage due to the available limitations can be used without any restrictions for the selection of the gas treating Units.

## 2. REFERENCES

Throughout this Standard the following standards and codes are referred to. The editions of these standards and codes that are in effect at the time of publication of this Standard shall, to the extent specified herein, form a part of this Standard. The applicability of changes in standards and codes that occur after the date of this Standard shall be mutually agreed upon by the Company and the Vendor/Consultant:

### IPS (IRANIAN PETROLEUM STANDARDS)

IPS-E-PR-190 "Layout and Spacing"

### ASTM (AMERICAN SOCIETY FOR TESTING AND MATERIALS)

ASTM A 516, Grade B  
ASTM A 106, Grade B  
ASTM A 333  
ASTM A 234, Grade WPB  
ASTM A 105, Grade LF<sub>2</sub>  
ASTM A 350, Grade LF<sub>2</sub>

### NACE (NATIONAL ASSOCIATION OF CORROSION ENGINEERS)

NACE Standard MR-01-75 "Material Requirements-Sulphide Stress Cracking Resistant Material for Oilfield Equipment", latest Revision

### ANSI (AMERICAN NATIONAL STANDARD INSTITUTE)

ANSI B 16.20

### TEMA (TUBULAR EXCHANGER MANUFACTURERS ASSOCIATION)

TEMA "Class R" Exchanger, TEMA "Class C" Exchanger, TEMA "Class B" Exchanger  
Standards of the Tubular Exchanger Manufacturers

**GPSA (GAS PROCESSORS SUPPLIERS ASSOCIATION)**

Vol. II, Ed. 10, 1987

**3. DEFINITIONS AND TERMINOLOGY****3.1 Absorption**

Absorption is a separation process involving the transfer of a substance from a gaseous phase to liquid phase through the phase boundary.

**3.2 Acid Gases**

Acid gases are impurities in a gas stream usually consisting of  $\text{CO}_2$ ,  $\text{H}_2\text{S}$ ,  $\text{COS}$ ,  $\text{RSH}$  and  $\text{SO}_2$ . Most common in natural gas are  $\text{CO}_2$ ,  $\text{H}_2\text{S}$  and  $\text{COS}$ .

**3.3 Acid Gas Loading**

Acid gas loading is the amount of acid gas, on a molar or volumetric basis, which will be picked up by a solvent.

**3.4 Adsorption**

Adsorption is a separation process involving the removal of a substance from a gas stream by physical binding on the surface of a solid material.

**3.5 Antifoam**

Antifoam is a substance, usually a silicone or long-chain alcohol, added to the treating system to reduce the tendency to foam.

**3.6 Degradation Products**

Degradation products are impurities in a treating solution which are formed both reversible and irreversible side reactions.

**3.7 Liquefied Petroleum Gas (LPG)**

Any material having a vapor pressure not exceeding that allowed for commercial propane composed predominantly of the following hydrocarbons, either by themselves or as a mixtures: propane, propylene, butane (normal butane or isobutane) and butylene, as a by-product in petroleum refining or natural gasoline manufacture.

**3.8 Mercaptan**

Mercaptan is a hydrocarbon group (usually a methane, ethane, or propane) with a sulfur group ( $-\text{SH}$ ) substituted on a terminal carbon atom.

**3.9 Normal cubic meters ( $\text{Nm}^3$ )**

Refers to capacity at Normal Conditions (101.325 kPa &  $0^\circ\text{C}$ ) and relative humidity of 0 percent.

### **3.10 Pressure Relief Valve**

A type of pressure relief device designed to both "open" and "close" to maintain internal fluid pressure.

### **3.11 Residence Time**

Residence time is the time period in which a fluid will be contained within a certain volume.

### **3.12 Selective Treating**

Selective treating is preferential removal of one acid gas component, leaving other acid components in the treated gas stream.

### **3.13 Sour Gas**

Sour gas is any gas stream which contains acid gas components.

### **3.14 Steric Hinderance**

Steric hinderance is chemically attaching a bulk molecule (such as benzene) to the hydrocarbon chain of an amine to inhibit CO<sub>2</sub> reacting to form a carbamate.

### **3.15 Sweet Gas**

Sweet gas is a gas stream which has acid gas components removed to an acceptable level.

### **3.16 Threshold Limit Value (TLV)**

Threshold limit value is the amount of a contaminant to which a person can have repeated exposure for an 8 hour day without adverse effects.

## **4. SYMBOLS AND ABBREVIATIONS**

<b>ANSI</b>	American National Standard Institute.
<b>ASTM</b>	American Society for Testing and Materials.
<b>CO<sub>2</sub></b>	Carbon Dioxide.
<b>CS<sub>2</sub></b>	Carbon Disulfide.
<b>DCS</b>	Distributed Control System.
<b>DEA</b>	di-Ethanolamine.
<b>DETA</b>	di-Ethylamine.
<b>DIPAM</b>	di-Isopropylamine.
<b>DN</b>	Diameter Nominal, in (mm).
<b>DGA</b>	di-Glycolamine.
<b>ESD</b>	Emergency Shut-Down.
<b>FOB</b>	Free On Board.
<b>GPSA</b>	Gas Processors Suppliers Association.
<b>HCN</b>	Hydrogen Cyanide.
<b>H<sub>2</sub>S</b>	Hydrogen Sulfide.
<b>HP</b>	High Pressure.
<b>ID</b>	Inside Diameter, in (mm).
<b>L</b>	liter.

<b>LP</b>	Low Pressure.
<b>LPG</b>	Liquefied Petroleum Gas.
<b>MEA</b>	mono-Ethanolamine.
<b>MDEL</b>	Methyl-di-Ethanolamine.
<b>max.</b>	maximum.
<b>min.</b>	minimum.
<b>NACE</b>	National Association of Corrosion Engineers.
<b>OD</b>	Outside Diameter, in (mm).
<b>Oper.</b>	Operating.
<b>OVHD</b>	Overhead.
<b>ppm<sub>m</sub></b>	parts per million by mass, in (mg/kg).
<b>ppm<sub>v</sub></b>	parts per million by volume, in ( $\mu\text{m}^3/\text{m}^3$ ).
<b>SS</b>	Stainless Steel.
<b>TEMA</b>	Tubular Exchanger Manufacturers Association.
<b>TLV</b>	Threshold Limit Value.
<b>WC</b>	Water Column, in (mm).
<b><math>\rho(\text{rho})</math></b>	Gas density at operating conditions, in ( $\text{kg}/\text{m}^3$ ).

## 5. UNITS

This Standard is based on International System of Units (SI), except where otherwise specified.

## PART I

### 6. GAS SWEETENING UNITS

#### 6.1 General

Acid gas constituents present in most natural gas streams are hydrogen sulfide ( $H_2S$ ) and carbon dioxide ( $CO_2$ ). Many gas streams, however, particularly those in a refinery or manufactured gases, may contain mercaptans, carbon sulfide or carbonyl sulfide.

The level of acid gas concentration in the sour gas is an important consideration for selecting the proper sweetening process. Some processes are applicable for removal of large quantities of acid gas, and other processes have the capacity for removing acid gas constituents to the parts per million (ppm) range. However, the sweetening process should meet the Pipeline Specification.

#### 6.2 Process Selection

##### 6.2.1 General

**6.2.1.1** The selection of a solvent process depends on process objectives and characteristics of the solvents, such as selectivity for  $H_2S$ , COS, HCN, etc., ease of handling water content in feed gas, ease of controlling water content of circulating solvent, concurrent hydrocarbon loss or removal with acid gas removal, costs, solvent supply, chemical inertness, royalty cost, thermal stability and proper plant performance for various processing techniques.

**6.2.1.2** The choice of the process solution is determined by the pressure and temperature conditions at which the gas to be treated is available, its composition with respect to major and minor constituents, and the purity requirements of the treated gas. In addition, consideration should be given to whether simultaneous  $H_2S$  and  $CO_2$  removal or selective  $H_2S$  absorption is desired.

**6.2.1.3** In general the gas treating process can affect the design of entire gas processing facilities including the methods used for acid gas disposal and sulfur recovery dehydration, absorbent recovery, etc.

**6.2.1.4** For an appropriate process selection, the following factors should be considered for evaluation and decision making as a general approach to all sour gas sweetening treatment installations:

- air pollution regulations regarding  $H_2S$  removal;
- type and concentration of impurities in sour gas;
- specification of treated gas (lean gas);
- temperature and pressure at which the sour gas is available and at which the sweetened gas should be delivered;
- volume of the gas to be treated;
- hydrocarbon composition of sour gas;
- selectivity required for acid gas removal;
- capital cost and operating cost;
- liquid product specifications.

**6.2.1.5** An accurate analysis of the sour gas stream should be made by the Company showing all of the impurities including COS,  $CO_2$  and mercaptans should be included in the project specification. These impurities shall in particular have significant effect on process design of the gas treating and down stream facilities.



**6.2.1.6** The selectivity of a sweetening agent is an indication of the degree of removal that can be obtained for one acid gas constituent as opposed to another. There are sweetening processes which display rather marked selectivity for one acid gas constituent. There are cases where no selectivity is demonstrated and all acid gas constituents will be removed.

**6.2.1.7** The selectivity of sweetening agent may be made on the basis of operating conditions. Only rarely will natural gas streams be sweetened at low pressures. However there are processes which are unsuitable for removing acid gases under low pressure conditions. Other sweetening agents may adversely be effected by temperatures, and some lose their economic advantage when large volumes of gas are to be treated.

The feasibility and desirability of sulfur recovery can also place considerable limitation on the selectivity of sweetening process. Removal of  $H_2S$  to produce contractually sweetened gas is a difficult thing and emphasis should be placed on those processes which have major potential application for sweetening of natural gases.

## **6.2.2 Chemical reaction processes**

### **6.2.2.1 General**

**6.2.2.1.1** Chemical reaction processes remove the  $H_2S$  and/or  $CO_2$  from the gas stream by chemical reaction with a material in the solvent solution. The alkanolamines are the most generally accepted and widely used of the many available solvents for removal of  $H_2S$  and  $CO_2$  from natural gas streams due to their reactivity and availability at low cost.

**6.2.2.1.2** The alkanolamine processes are particularly applicable where acid gas partial pressures are low or low levels of acid gas are considered in the residue gas. Physical properties for some of the gas treating chemicals are shown in Table 1. The alkanolamines are clear, colorless liquids that have a slightly pungent odor. All except tri-Ethanolamine are considered stable materials, because they can be heated to their boiling points without decomposition. Tri-Ethanolamine decomposes at below its normal boiling point.

TABLE 1 - PHYSICAL PROPERTIES OF GAS TREATING CHEMICALS

	Monothalamine	Diethalamine	Triethalamine	Diglydamine	Dioctalamine	Diethylamine	Diethylamine
Formula	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$	$\text{HOC}_2\text{H}_4\text{NH}_2$
Molecular mass	61.08	103.14	148.19	105.14	133.19	133.19	133.19
Boiling point @ 101.3 kPa (abs), °C	170.3	269	360 (Decomposes)	271	249	270	270
Freezing point, °C	10.5	28	22.4	-12.5	42	-28.9	-40.2
Critical constants							
Pressure, kPa (abs)	5985	3273	7448	3772	3770	—	—
Temperature, °C	330	442	514	403	109	—	—
Density @ 20°C, kg/m <sup>3</sup>	1018	1093	1124	1038 @ 15.6°C	999 @ 30°C	1031 @ 25°C	1205.7
Relative density, 20°C/20°C	1.0179	1.0019 (30/30°C)	1.1258	1.0572	0.989 @ 45/20°C	—	1.203
Specific heat capacity, kJ/(kg °C) @ 15.6°C	2.366 @ 20°C	2.512	2.931	2.391	2.889 @ 30°C	2.032 @ 5°C	1.403
Thermal conductivity, J/(s·m <sup>2</sup> ·°C/m) @ 20°C	0.236	0.320	—	0.209	—	0.190 @ 25°C	0.208 @ 17°C
Latent heat of vaporization, kJ/kg at kPa (abs)	826 @ 101.3 kPa	670 @ 9.73 kPa	533 @ 101.3 kPa	510 @ 101.3 kPa	430 @ 101.3 kPa	—	484 @ 101.3 kPa
Heat of reaction, kJ/kg of acid gas							
H <sub>2</sub> S	—	—	-920	-1348	—	-442 @ 25°C	—
CO <sub>2</sub>	—	—	-1463	-1977	—	-372 @ 25°C	—
Viscosity, mPa s	24.1 @ 20°C	350 @ 20°C	1013 @ 20°C	40 @ 15.6°C	198 @ 45°C	5.8 @ 25°C	10.8 @ -23°C
		80 mass % solution	85 mass % solution		86 @ 54°C		4.5 @ 0°C
							2.5 @ 25°C
							0.87 @ 100°C
							1.4209
Refractive index, n <sub>D</sub> , 20°C	1.4339	1.4776	1.4852	1.4598	1.4547 @ 45°C	—	—
Flash point, °C	93	138	183	127	124	151	137

	Diethylene Glycol	Triethylene Glycol	Tetraethylene Glycol	Pentaethylene Glycol	Hexaethylene Glycol	Diethylene Glycol	Diethylene Glycol
Formula	$\text{HOC}_2\text{H}_4\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OCH}_2\text{CH}_2\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OH}$	$\text{HOC}_2\text{H}_4\text{OCH}_2\text{CH}_2\text{OH}$
Molecular mass	62.07	197	334	474	620	120.17	120.17
Boiling point @ 101.3 kPa (abs), °C	197	245	357	474	620	285	285
Freezing point, °C	-13	-8	-13	-13	-13	27.6	27.6
Critical constants							
Pressure, kPa (abs)	7696	4661	4661	4661	4661	5290	5290
Temperature, °C	372	408	408	408	408	543	543
Density @ 20°C, kg/m <sup>3</sup>	1113	1113	1113	1113	1113	1283	1283
Relative density, 20°C/20°C	1.1135	1.1184	1.1254	1.1254	1.1254	1.268	1.268
Specific heat capacity, kJ/(kg °C) @ 15.6°C	2.311	2.340	2.340	2.340	2.340	1.507 @ 30°C	1.507 @ 30°C
Thermal conductivity, J/(s·m <sup>2</sup> ·°C/m) @ 20°C	0.287 @ 15.6°C	0.287 @ 15.6°C	0.287 @ 15.6°C	0.287 @ 15.6°C	0.287 @ 15.6°C	0.197 @ 37.8°C	0.197 @ 37.8°C
Latent heat of vaporization, kJ/kg at kPa (abs)	800 @ 101.3 kPa	540 @ 101.3 kPa	416 @ 101.3 kPa	416 @ 101.3 kPa	416 @ 101.3 kPa	325 @ 100°C	325 @ 100°C
Viscosity, mPa s	20.9 @ 20°C	35.7 @ 20°C	47.9 @ 20°C	47.9 @ 20°C	47.9 @ 20°C	10.3 @ 30°C	10.3 @ 30°C
						6.1 @ 50°C	6.1 @ 50°C
						2.5 @ 100°C	2.5 @ 100°C
						1.4 @ 150°C	1.4 @ 150°C
						0.97 @ 200°C	0.97 @ 200°C
						1.481 @ 30°C	1.481 @ 30°C
						177	177
						1.3286	1.3286
						14	14

NOTE: 1 mPa s = 1 centipoise

### 6.2.2.2 mono-Ethanolamine (MEA)

**6.2.2.2.1** MEA should commonly be used as a 10 to 20% solution in water. The acid gas loading should usually be limited to 0.3 to 0.4 moles acid gas per mole of amine for carbon steel equipment.

**6.2.2.2.2** MEA itself is not considered to be particularly corrosive. However, its degradation products are very corrosive. COS, CS<sub>2</sub>, SO<sub>2</sub> and SO<sub>3</sub> can partially deactivate MEA, which may essentially require to be recovered with a reclaimer.

**6.2.2.2.3** Since MEA is primary amine, it has a high pH. This enables MEA solutions to produce gas containing less than 6 mg/Sm<sup>3</sup> (¼ grains H<sub>2</sub>S per 100 Scu.ft) of acid gas at very low H<sub>2</sub>S partial pressures.

**6.2.2.2.4** The heat of reaction for CO<sub>2</sub> in MEA is about 1930 kJ/kg of CO<sub>2</sub> (460 kcal/kg of CO<sub>2</sub>). The heat of reaction for all amines is a function of loading and other conditions. It varies by only 117 to 138 kJ/kg (28 to 33 kcal/kg) up to about 0.5 mole/mole of total acid gas loadings. Above this loading, the heat of reaction varies considerably and should be calculated as a function of loading.

**6.2.2.2.5** MEA will easily reduce acid gas concentrations to Pipeline Specifications (generally less than 6 mg H<sub>2</sub>S/Sm<sup>3</sup> gas (0.25 grains per 100 Scu.ft). By proper design and operation, the acid gas content can be reduced as low as 1.2 mg H<sub>2</sub>S/Sm<sup>3</sup> gas (0.05 grains per 100 Scu.ft).

### 6.2.2.3 di-Ethanolamine (DEA)

**6.2.2.3.1** DEA is commonly used in the 25 to 35 mass percent range. The loading for DEA is also limited to 0.3 to 0.4 mole/mole of acid gas for carbon steel equipment.

**6.2.2.3.2** When using stainless steel equipment, DEA can safely be loaded to equilibrium. This condition can be considered for carbon steel equipment by adding inhibitors.

**6.2.2.3.3** The degradation products of DEA are much less corrosive than those of MEA. COS and CS<sub>2</sub> may irreversibly react with DEA to some extent.

**6.2.2.3.4** Since DEA is a secondary alkanolamine, it has a reduced affinity for H<sub>2</sub>S and CO<sub>2</sub>. As a result, for some low pressure gas streams, DEA can not produce Pipeline Specification gas.

**6.2.2.3.5** Under some conditions, such as low pressure and liquid residence time on the tray (of about 2 seconds), DEA will be selective toward H<sub>2</sub>S and will permit a significant fraction of CO<sub>2</sub> to remain in the product gas.

**6.2.2.3.6** The heat of reaction for DEA and CO<sub>2</sub> is 151 kJ/kg of CO<sub>2</sub> (360 kcal/kg of CO<sub>2</sub>) which is about 22% less than for MEA.

### 6.2.2.4 di-Glycolamine (DGA)

**6.2.2.4.1** DGA is generally used as 40 to 60 mass percent solutions in water. The reduced corrosion problems with DGA allow mole per mole solution loadings equivalent to MEA in most applications even with these high mass percentage.

**6.2.2.4.2** For gas streams with acid gas partial pressures, absorber bottoms temperatures as high as 82°C and above can occur. This will reduce the possible loading.

**6.2.2.4.3** DGA has a tendency to preferentially react with CO<sub>2</sub> over H<sub>2</sub>S. It also has a higher pH than MEA and thus can easily achieve 6 mg H<sub>2</sub>S/ Sm<sup>3</sup> gas (0.25 grains per 100 Scu.ft) except in some cases where large amounts of CO<sub>2</sub> are present relative to H<sub>2</sub>S.

**6.2.2.4.4** DGA has some definite advantages over the other amines in that higher DGA concentrations in the solution result in lower freezing points. One of the primary advantages of DGA is high heats of reaction as compared with other amines and shown in the following Table 2 or comparative data Table.

TABLE 2 - COMPARATIVE DATA TABLE

AMINE	MEA	DEA	DGA	MDEA
Solution strength, mass %	15-20	25-35	40-60	30-50
Acid gas loading mole/mole	0.3-0.4	0.3-0.4	0.3-0.4	Unlimited
$\Delta H^*_{\text{r}}$ for $\text{H}_2\text{S}$ kJ/kg (kcal/kg)	1281 (306)	1189 (284)	1570 (375)	1214 (290)
$\Delta H^*_{\text{r}}$ for $\text{CO}_2$ kJ/kg (kcal/kg)	1922 (459)	1520 (363)	1729 (413)	1398 (334)
Ability to preferentially absorb $\text{H}_2\text{S}$	No	Under some conditions	No	Under most conditions

\* kJ/kg (kcal/kg) and for total loadings below 0.5 moles acid gas/mole amine

#### 6.2.2.5 Methyl di-Ethanolamine (MDEA)

**6.2.2.5.1** MDEA is most commonly used in the 30 to 50 mass percent range. Due to considerably reduced corrosion problems, acid gas loadings as high as 0.7 to 0.8 mole/mole are practical in carbon steel equipment.

**6.2.2.5.2** Since MDEA is a tertiary amine, it has less affinity for  $\text{H}_2\text{S}$  and  $\text{CO}_2$  than DEA. Thus, as in the case for DEA, MDEA can not produce Pipeline Specification gas for some low pressure streams.

**6.2.2.5.3** MDEA has several distinct advantages over primary and secondary amines. These include:

Lower vapor pressure, lower heats of reaction, higher resistance to degradation, fewer corrosion problems and selectivity toward  $\text{H}_2\text{S}$  in the presence of  $\text{CO}_2$ .

#### 6.2.2.6 General considerations

**6.2.2.6.1** Over the years of operation of amine sweetening Units a substantial amount of information has been reported, which deem necessary to be considered in the design stages of a sweetening Units as following:

##### a) Corrosion

**a.1)** The temperature of the solution in the reboiler and the temperature of the steam should be kept as low as possible.

**a.2)** The use of high-temperature heat-carrying media, such as oil in reboilers should be avoided to maintain the lowest possible metal skin temperature.

**a.3)** Lowest possible pressure on stripping columns and reboilers should be considered to avoid severe corrosion of reboiler tubes.

**a.4)** Inert gas blanketing facilities should be considered over all portion of the solution which could be exposed to the atmosphere. A positive pressure on the suction side of the pumps should also be ensured for, excluding oxygen from the system.

##### b) Inlet scrubbing

The importance of proper and efficient inlet scrubbing of the sour gas is difficult to over-emphasize. Many problems in amine plant operations such as foaming, corrosion and reboiler tube burn-out can be traced to the presence of excessive amounts of foreign material in the amine solution. Liquid hydrocarbons and entrained solids

frequently can enter the plant with the sour gas stream. In addition, such foreign materials as corrosion inhibitors, drilling muds and well acidizers can also enter with the sour gas stream, particularly on lease-type sweetening Units. The inlet separation equipment should be sized and designed with these factors in mind. The fact that these materials do not enter at a steady rate but slug or surge into the plant should also be kept in mind. Instantaneous flow rates may be extremely high and the inlet separating system must be designed to handle that.

### **c) Amine losses**

Amine losses can be very expensive. A separator on the sweet gas stream leaving the contactor is advisable. This, will also help to eliminate amine losses from unexpected foaming or surges. The savings in down time and loss of production will be other advantages of the sweet gas scrubber, which may be many times of the savings in amine solution lost.

### **d) Filtration**

**d.1)** Proper solution is important for the maintenance of a clean, efficient amine solution. Positive removal of all solids from the system minimizes corrosion and associated problems.

In general, a two-stage filtration is recommended often for this service. Any secondary filtration may be activated charcoal and will remove degradation products, should they exist, along with some of the smaller particles of entrained solids. The filter system should be capable of handling at least 10 to 20% of the amine circulation rate. Doing this will permit quick clean up of the amine solution after an upset.

**d.2)** With full-flow filtration, parallel filters with no by-pass is recommended. The best place for filter location may be considered (i) on the rich amine solution at the outlet of the contactor, (ii) on the lean amine solution just before the solution enters the contactor.

### **e) Amine reclaiming**

**e.1)** The entrained solids, desolved salts and degradation products that cause foaming and corrosion problems is removed by reclaimer. The following points should be considered for reclaimer:

- i)** An easy-to-open manway should be located on the reclaimer shell so the solids can be simply washed out at the end of the cycle. The drain line should be large enough to pass solids.
- ii)** The tube bundles should be raised 15 cm or more from the bottom of the reclaimer shell to provide space for sludge accumulation below the tubes, and to give better solution flow around the tubes.
- iii)** A packed column should be placed on top of the reclaimer to eliminate foam and entrainment from the overhead vapor stream. The installation of glass site port in the vapor line will help keep a check on this carry over. An analytical evaluation will help pinpoint suspected entrainment.
- iv)** The tubes should be widely spaced for easy cleaning.
- v)** Make certain the steam supply is not superheated.
- vi)** A chart recorder should be used to monitor the reclaimer temperature throughout the cycle.
- vii)** Amine feed to the reclaimer should be controlled by a level controller on the kettle to maintain a liquid level at least 15 cm above the tube. A rotameter should be installed in the feed line to indicate when the feed falls off and steam stripping should be started.
- viii)** Sufficient vapor space should be allowed above the liquid layer in the reclaimer to prevent liquid carryover in the overhead vapor line.
- ix)** Provisions for chemical analysis of the reclaimer bottoms and the amine solution should be considered to identify the solution contaminants and determine the degree and rate of solution contamination.

#### **f) Amine-amine heat exchanger**

If an intermediate flash separator is not used, contactor pressure should be maintained through the amine-to-amine heat exchanger. Doing this minimizes breakout of acid gases from the rich amine solution, thereby minimizing excessive corrosion of control valves, heat exchanger and downstream piping. Operation at high pressure increases the cost of the exchangers, but this generally will be more than offset by the increased life due to lessened corrosion. The rich amine solution should be on the tube side of the exchanger, which again tends to minimize investment in the exchanger.

Linear velocities in the amine-to-amine heat exchanger should be low-in the range of 0.6 to 1.0 meter per second. This reduces the heat transfer coefficient and increases the surface area requirement. However, again the return on the reduced maintenance costs for the exchanger will more than offset the increased investment. Flowing amine solutions should not impinge directly on vessel surfaces; impingement baffles should be utilized in the exchangers.

#### **g) Reboiler of regenerator**

For careful and accurate design of the reboiler system in amine regeneration section the following points should be considered:

- g.1)** Low pressure saturated steam at approximately 300-380 kPa (135-142°C) is commonly used to strip the amine solution. Steam temperatures above 140°C should be avoided to prevent excessive skin temperatures on the tubes.
- g.2)** Amine Units should be designed for "pressure operation". Higher operating pressures increase the bottom temperature of the regenerator and are believed to provide more complete stripping of the acid gases, especially the CO<sub>2</sub>.
- g.3)** The maximum allowable kettle temperature recommended for regeneration of the solution is 127°C to prevent amine degradation.
- g.4)** The temperature controller valve should be on the steam inlet, not on the condensate outlet, to keep excessive condensate out of the tubes. When condensate partially floods the reboiler tube, the heat load concentrates in the top section of the bundle which can cause tube failure.
- g.5)** Stripping steam requirements will vary depending on the degree of sweetening required for the process stream being treated. Normally, the steam consumption should be at least 120 grams of excess steam per liter of solution circulated.
- g.6)** To provide good circulation of the amine solution around the tubes and to reduce fouling caused by a sludge accumulation, the tube bundle should be placed on a slide about 15 cm above the bottom of the reboiler shell.
- g.7)** The amine solution should enter the reboiler at several locations to help improve the natural circulation of liquid in the reboiler shell. Several vapor exit locations will reduce the stagnant pockets of acid gases in the reboiler.
- g.8)** The tube bundle should be supported to prevent tube vibration. Teflon or other protective inserts will prevent tube cutting.
- g.9)** The length of the tubes should be limited to prevent condensate "logging" and water hammer. The water hammer effect can produce severe tube vibration and grooving of the tubes at the support baffles.
- g.10)** A square pitch tube pattern is recommended for both reboiler and heat exchange bundles to provide easy cleaning. The tubes should be widely spaced to permit a rapid escape of liberated gases and to reduce the high velocity scrubbing action associated with two phase flow. This scrubbing action will remove the protective film and increase corrosion.

**g.11)** The reboiler should be designed to provide a liberal amount of vapor disengaging space between tubes, and with sufficient surface to produce a "simmering" action rather than violent boiling. In existing installations where vapor binding is a problem, some tubes can be removed to form an "X" or "V" in the center of the bundle to provide a path of low resistance to escaping vapors.

**g.12)** The reboiler bundle should always be kept covered with 15-20 centimeter of liquid to prevent localized drying and overheating. Severe corrosion will occur if the liquid level is lowered until some of the tubes are exposed.

**g.13)** An analysis of the amine solution entering and leaving the reboiler will determine the efficiency of the stripping operation. A high acid gas loading in the reboiler causes tube corrosion.

### 6.2.2.7 Summary and conclusions

Many process factors should be considered when selecting an amine for a sweetening application as outlined hereinunder:

- a)** The initial selection should be based on the pressure and acid gas content of the sour gas as well as the purity specification for the product gas.
- b)** Based on the currently "accepted" operating conditions, MEA is usually not preferred for its high heat of reaction and lower acid gas carrying capacity per unit volume of solution. However, MEA is still used for plants where the inlet gas pressure is low and Pipeline Specification gas or total removal of the acid gases are required.
- c)** DEA, is used for its lower heats of reaction, higher acid gas carrying capacity and resultant lower energy requirements. However, its potential for selective H<sub>2</sub>S removal from streams containing CO<sub>2</sub> has not fully been realized.
- d)** DGA, although has high heat of reaction, its very high gas carrying capacity usually produces very reasonable net energy requirements. DGA also has a good potential for absorbing COS and some mercaptans from gas and liquid streams, and because of this, DGA has been used in both natural and refinery gas applications.
- e)** MDEA, with its some outstanding capabilities, resulting from its low heat of reaction, can be used in pressure swing plants for bulk acid gas removal. MDEA is currently best known for its ability to preferentially absorb H<sub>2</sub>S.

### 6.2.3 Combined physical/chemical purification processes

**6.2.3.1** The newly classified physical/chemical combined purification process i.e., alkanol amines (mono-or di-ethanol amine) mixed with methanol are more successful than a single physical solvent used in many of the gas treatment plants. The main advantage of this solvent lies in the good physical absorption of physical solvent component in combination with the chemical reaction of the amine. The combination of a chemically active amine with a low boiling point polar physical solvent such as methanol offers major advantages in the absorption of CO<sub>2</sub> and sulfur components as:

- a)** Very low clean gas sulfur contents of less than 0.1 ppm, as they are required for synthesis gases.
- b)** Very low CO<sub>2</sub> contents in the purified gas.
- c)** Good absorption of trace components, such as HCN, COS, mercaptans and higher hydrocarbons.
- d)** Low regeneration temperature, owing to the fact that the solvent methanol has a boiling point approximately 35°C below that of water.
- e)** The solvent is non-corrosive so that carbon steel equipment can be used.

**6.2.3.2** As the conventional alkanolamines such as mono-Ethanol Amine (MEA) and di-Ethanol Amine (DEA) are used as non-selective solvents, the sulfur components H<sub>2</sub>S and COS are absorbed together with CO<sub>2</sub> and jointly occur in

the off-gas. This is a disadvantage, as the sulfur rich off-gas, should be treated. These disadvantages have been eliminated by an extensive research and MEA/DEA are replaced with following aliphatic alkylamines:

- 1) di-Isopropylamine      - DIPAM.
- 2) di-Ethylamine         - DETA

**6.2.3.3** These new DIPAM and DETA amines offer the following advantages against the conventional MEA and DEA:

- a) High thermal and chemical stability so that no reclaimer is needed.
- b) Higher effective CO<sub>2</sub> and H<sub>2</sub>S loadings.
- c) High selectivity of the solvent between sulfur components and CO<sub>2</sub>, along with the very low residual sulfur contents in the clean gas.
- d) Good industrial availability of these amines at moderate cost.
- e) Smaller difference between absorption and desorption temperature.
- f) No foaming tendency due to low surface tension.
- g) No corrosion problems.
- h) Considerable reduction in vapor pressure of the amine neutralized by sour gases.

**6.2.3.4** A particular advantage of the aliphatic alkylamines, is the high selectivity of the solvent between sulfur components and CO<sub>2</sub>, along with very low residual sulfur contents in the clean gas. Even with very low sulfur and high CO<sub>2</sub> contents in feed gas, it is unproblematic to produce a high sulfur off-gas containing more than 25% H<sub>2</sub>S, which is suitable for Claus plants.

**6.2.3.5** The CO<sub>2</sub> produced by alkylamine process is free from COS and shall meet today's stringent environmental pollution requirements. CO<sub>2</sub> produced in this process is pure and can be used for urea synthesis.

**6.2.3.6** Taking the advantage of varied properties of aliphatic alkylamine, the process lends itself mainly for the following applications:

- a) Joint absorption of H<sub>2</sub>S, COS and CO<sub>2</sub> as well as trace contaminants, such as HCN, NH<sub>3</sub>, mercaptans, thiophenes, etc., from raw gases to produce highly pure synthesis gases.
- b) Selective absorption of all sulfur components from raw gases to obtain highly pure product gas and a high sulfur Claus gas. CO<sub>2</sub> is removed in second stage and can be recovered in pure form.

**6.2.3.7** It should be taken in mind that, combined physical/chemical purification processes and aliphatic alkylamine processes are having predominantly chemical characteristics and so, these processes are particularly cost-effective for all gases with low to medium partial pressures of the sour gas and if the sour gas partial pressures are high, physical gas purification processes should preferably be used.

## **6.2.4 Carbonate process**

**6.2.4.1** The Hot Potassium Carbonate Process has been utilized successfully for bulk removal of CO<sub>2</sub> from a number of gas mixtures. It has been used for sweetening natural gases containing both CO<sub>2</sub> and H<sub>2</sub>S.

**6.2.4.2** Hot Potassium Carbonate Process is not suitable for sweetening gas mixtures containing little or not CO<sub>2</sub>, as the potassium bisulfide should be very difficult to regenerate if CO<sub>2</sub> is not present.



#### 6.2.4.3 This process has the following advantages/disadvantages:

##### a) Advantages;

- a.1) It is a continuous circulating system employing an inexpensive chemical.
- a.2) It is an iso-therm system in that both absorption and desorption of the acid gas are conducted at as nearly a uniform high temperature as can be obtained, thus requiring no heat exchange equipment in the fluid circulating system.
- a.3) The desorption by stripping is accomplished with a smaller steam rate than required for an amine plant.

##### b) Disadvantages;

- b.1) It will not commercially reduce the  $H_2S$  content to so-called Pipeline Specification. To accomplish this, a conventional amine plant should be used.
- b.2) It is similar to other acid gas removal processes in that it is also prone to corrosion.
- b.3) Like other liquid absorbents in sweetening plants, it is liable to the problems of suspended solids and foaming.

#### 6.2.4.4 Corrosion

**6.2.4.4.1** In carbonate process corrosion will occur where the conversion to bicarbonate is high and where carbon dioxide and steam are released by pressure reduction. In addition, stress corrosion is expected because potassium carbonate is an electrolyte. This requires stress relieving of welds and vessels.

**6.2.4.4.2** Successful corrosion inhibitors that so far reported includes arsenic and vanadium salts as well as dichromates.

**6.2.4.4.3** Pilot plant studies conducted by the Bureau of Mines encountered severe corrosion of carbonate in several areas. In general this corrosion occurred where the conversion to bicarbonate was occurred where carbon dioxide and steam were released by pressure reduction. In addition, stress corrosion expected because potassium carbonate is an electrolyte. This requires stress relieving of welded vessels.

**6.2.4.4.4** Successful corrosion inhibitors that have been reported in the literature include arsenic and vanadium salts as well as dichromates.

**6.2.4.4.5** Stainless steel is recommended for reboiler tubes, though carbon steel has been used successfully. Control valves and solution pumps should also be of stainless steel, particularly those parts which encounter the carbonate solution such as impellers and inner valves.

**6.2.4.4.6** The absorber and stripper, whether using trays or packing, can be made of carbon steel. They should be stress relieved. Piping can also be carbon steel if stress relieved. Stripper columns in particular are sometimes plastic coated or gunnite lined.

**6.2.4.4.7** When carbon dioxide is the only acid gas constituent, potassium dichromate will serve as an effective corrosion inhibitor. From 1,000 to 3,000 ppm will generally be effective. Where  $H_2S$  is present in the gas, chromates cannot be used. Both  $H_2S$  and COS cause reduction of the hexavalent chromium to the insoluble trivalent state. When this happens the corrosion inhibitor causes erosion and precipitation. Fortunately, if sufficient quantity of  $H_2S$  is present, formation of the potassium bisulfide inhibits corrosion to a significant extent. In cases where corrosion is a problem with  $H_2S$  present, the commercially available film-forming fatty amines are usually effective as corrosion inhibitors.

**6.2.4.4.8** Though water vapor exerts a partial pressure over solutions of potassium carbonate, the potassium carbonate does not. For this reason, in a process Unit with properly sized vessel and good mist extractors, potassium carbonate losses should be negligible. Products of undesirable side reactions, resulting from trace gas impurities, can be maintained at satisfactory levels by a small continuous purge of carbonate solution.

**6.2.4.4.9** The carbonate solution should be filtered through a side stream filter arrangement. There will be occasional pieces of scale and other solid materials that the filter should remove. Filtration minimizes severe erosion problems.

### **6.2.5 Catacarb process**

**6.2.5.1** The Catacarb Process employs a modified potassium salt solution containing a very active, stable and non-toxic catalyst and corrosion inhibitor. In this process amine borate is utilized to increase the activity of the hot potassium carbonate.

**6.2.5.2** Several other catalysts and inhibitors are used in this process. The choice depends on the gas compositions to be treated by the process. This process is also capable of removing trace amounts of other acid gases such as COS, CS<sub>2</sub> and RSH.

### **6.2.6 Physical absorption methods**

#### **6.2.6.1 General considerations**

**6.2.6.1.1** Various amines, Hot Potassium Carbonate and Catacarb Processes rely on the chemical reaction to remove acid gas constituent from sour gas streams. Removal of acid gases by physical absorption is possible and there are number of competitive processes which should be considered when:

- a) The partial pressure of the acid gas in the feed is greater than 350 kPa or 3.5 bar (50 psi). For this figure reference is made to GPSA in Clause 2.
- b) The heavy hydrocarbon concentration in the feed gas is low.
- c) Bulk removal of the acid gas is desired.
- d) Selective removal of H<sub>2</sub>S is desired.
- e) In general, physical solvents are capable of removing COS, CS<sub>2</sub> and mercaptans.

#### **6.2.6.2 Solvent regeneration**

**6.2.6.2.1** The solvents are regenerated by:

- a) multi-stage flushing to low pressure;
- b) regeneration at low temperature with an inert stripping gas;
- c) heating and stripping of solution with steam/solvent vapor.

#### **6.2.6.3 Flour process**

**6.2.6.3.1** The main characteristics desired in Flours Process for right selection of solvent are:

##### **a) Low vapor pressure at operating temperature**

This is desirable from two standpoints, both related. If solvent vapor pressure is appreciable, losses will be high. These losses necessitate either a complicated solvent recovery system or high operating costs due to solvent loss. For these reasons, Flour eliminated from their consideration several high vapor pressure solvents that had good solubilities for acid gas constituents.

**b) The primary constituents in the gas stream should be only slightly, if at all, soluble in the solvent**

In the case of a natural gas stream, methane and the heavier hydrocarbons should not be appreciably soluble in the solvent. If they are soluble, then expensive and complicated procedures may be required to prevent excessive losses.

**c) The solvent should have low viscosity**

High viscosities increase pumping costs. In general they also have an adverse effect on tray efficiencies and mass transfer. Operation at sub-ambient temperatures may aggravate the viscosity problem. For these cases some solvent satisfactory for ambient temperature operation might prove undesirable at lower temperature.

**d) Low solubility for water**

In general, increasing the water content in the circulating solvent will lower its carrying capacity for acid gases. Dissolved water will also tend to increase corrosion and solvent decomposition effects. If water is dissolved then steps must be taken to maintain the water content of the solvent at some specified level. Again, this increases plant complexity, costs and operational problems.

**e) The solvent should not degrade under normal operating conditions**

The solvent should not degrade chemically under the temperature and pressure conditions of normal operation. This problem can be handled by filtration, reclaiming and so forth, but these items do increase the investment and operating costs. The solvent should be stable with regard to oxygen and other materials. Again, storage tanks can be inert gas blanketed but this is a complicating factor to be avoided if at all possible.

**f) The solvent should not react chemically with any component in the gas stream**

This can lead to solvent degradation loss and loss of solvent effectiveness.

**g) For obvious reasons, the solvent should be non-corrosive to the common metals**

Use of carbon steel construction, preferably without necessity for stress relieving, will minimize plant investment. In the physical absorption process conditions are usually ideal for minimizing corrosion due to carbon dioxide and hydrogen sulfide. Temperatures are low and a proper solvent should help to minimize corrosion problems rather than cause them.

**h) The solvent should be readily available at reasonable cost**

An excellent solvent which, because of its cost, would have an appreciable effect on plant investment would not be desirable.

**6.2.6.3.2** Using the above characteristics as a guideline, the following solvents are recommended for use for the removal of carbon dioxide from high pressure gas streams:

- a)** Polyethylene carbonate.
- b)** Butoxy diethylene glycol acetate.
- c)** Glycol triacetate.
- d)** Methoxy triethylene glycol acetate.

#### 6.2.6.4 Other miscellaneous physical methods

**6.2.6.4.1** There are newly commercialized physical solvent processes which should be evaluated for proper process selection in sour gas treatment projects. Most of the equilibrium data of these solvents are proprietary to the process Licensors. Therefore, definitive comparative information about the solvent performance should be achieved in the course of process evaluation and License selection.

**6.2.6.4.2** Some of these physical solvent processes and their comparative data are given in Appendix A, Tables A.1 and A.2.

#### 6.2.7 Solid bed sweetening methods (batch processes)

##### 6.2.7.1 General

**6.2.7.1.1** The use of solids for sweetening gas is based on adsorption of the acid gases on the surface of the solid sweetening agent, or reaction with some component on that surface. These types of processes are not as widely used as the liquid processes but there are several advantages that make them worth considering.

**6.2.7.1.2** The solids processes are usually best applied to gases containing low-to-medium concentrations of  $H_2S$  or mercaptans. The solids processes tend to be highly selective and do not normally remove significant quantities of  $CO_2$ . Consequently, the  $H_2S$  stream from the process is usually high purity. In addition, pressure has relatively little effect on the adsorptive capacity of a sweetening agent. Most of solids processes are of the batch type and tend to have low investment and operating costs.

##### 6.2.7.2 Iron oxide (sponge) process

**6.2.7.2.1** The iron oxide or dry box process is one of the oldest known methods for removal of sulfur compounds from gas streams. It was introduced to England about the middle of the 19th Century and is still widely used in many areas of special application.

**6.2.7.2.2** This process often offers advantage when the sulfur in the gas does not exceed 7-9 tonnes per day and the concentration does not exceed 2400 g/100 Sm<sup>3</sup> [1000 grains  $H_2S$  per 100 Scf] of gas.

**6.2.7.2.3** This process consists of wood chips impregnated with varying amounts of hydrate iron oxide ( $Fe_2O_3$ ). This reacts with the  $H_2S$  to form  $Fe_2S_3$ , which may be regenerated with air. Continuous regeneration is possible by injecting a small stream of air into the feed-gas stream which converts the sulfide to the oxide and liberates the elemental sulfur. Regeneration is normally finished when the outlet oxygen concentration reaches 4-6% and the bed outlet temperature starts dropping. Each charge of sponge may be regenerated several times, but it gradually becomes less efficient and requires replacement.

##### 6.2.7.2.4 Advantages of iron sponge process

- a) Providing complete removal of small to medium concentrations of hydrogen sulfide without removing carbon dioxide.
- b) Requiring relatively small investment, for small to moderate gas volumes, compared with other processes.
- c) Being equally effective at any operating pressure.
- d) Being used to remove mercaptans or convert them to disulfides.

##### 6.2.7.2.5 Disadvantages of the iron sponge process

The disadvantages of the process are:

- a) It is a batch process, requiring duplicate installation or flow interruption of processed gas.

- b) It is prone to hydrate formation when operated at higher pressures and at temperatures in the hydrate-forming range.
- c) It will effectually remove ethyl mercaptane that has been added for odorization.
- d) Coating of the iron sponge with entrained oil or distillate will require more frequent change out of the sponge bed.

### 6.2.7.3 Molecular sieves

#### 6.2.7.3.1 General

Crystalline sodium-calcium aluminosilicates can be used for selective removal of  $\text{H}_2\text{S}$  and other sulfur compounds from natural gas streams. Some crystalline forms of these materials are found naturally. The common crystalline forms used in commercial adsorption are synthetically manufactured materials. The activated crystalline material is porous. The pore openings in a given structural are all exactly the same size and are determined by the molecular structure of the crystal and the size of molecules present in the crystal. The pores are formed by driving off water of crystallization that is present during the synthesis process. The exactness of the pore size and distribution has given rise to the name molecular sieves, which is used almost universally to describe these materials.

Molecular sieves have the large surface area typical of any solid adsorbent. In addition, however, molecular sieves have highly localized polar charges. These localized charges are the reason for the very strong adsorption of polar or polarizable compounds on molecular sieves. This also results in much higher adsorptive capacities for these materials by molecular sieves than by other adsorbents, particularly in the lower concentration ranges.

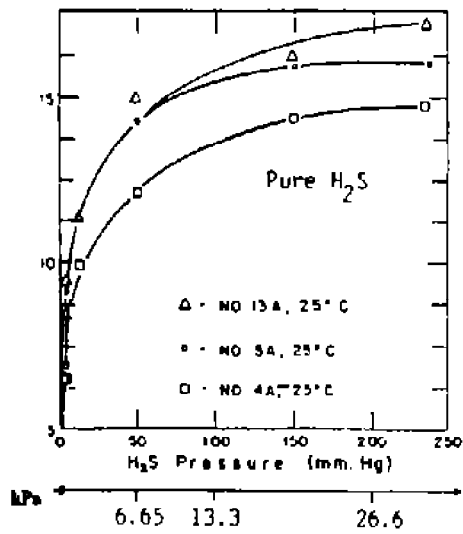
#### 6.2.7.3.2 Basic capacity and performance

Figs. 1 and 2 illustrate one interesting characteristic of molecular sieves. Fig. 1 shows that the adsorptive capacity in pure  $\text{H}_2\text{S}$  gas approaches about 0.15 kg (pounds) of  $\text{H}_2\text{S}$  per kg (pounds) of mole sieve under static conditions and at equilibrium saturation. Fig. 2 shows that there is a much lower saturation level when the  $\text{H}_2\text{S}$  is in a mixture rather than as a pure component.

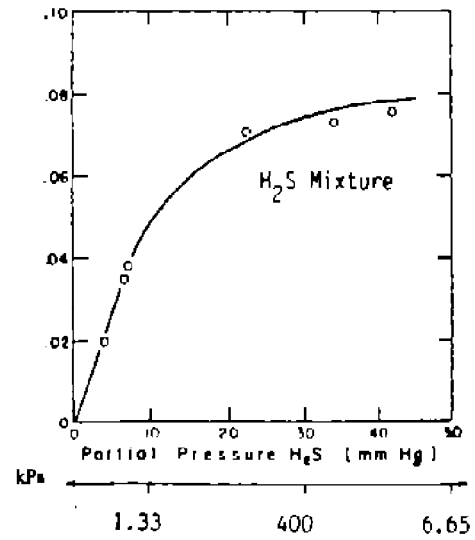
Fig. 3 shows that the adsorptive capacity of mole sieves for  $\text{H}_2\text{S}$  increases as temperature is lowered. Also shown is the effect of  $\text{CO}_2/\text{H}_2\text{S}$  ratio. As the percentage  $\text{H}_2\text{S}$  in the mixture increases the adsorptive capacity of the molecular sieve for  $\text{H}_2\text{S}$  increases.

Fig. 4 shows the effect of contact time on adsorptive capacity. Beyond approximately 170 seconds, increasing contact time has no effect on the capacity.

kg H<sub>2</sub>S Adsorbed/kg mol sieve

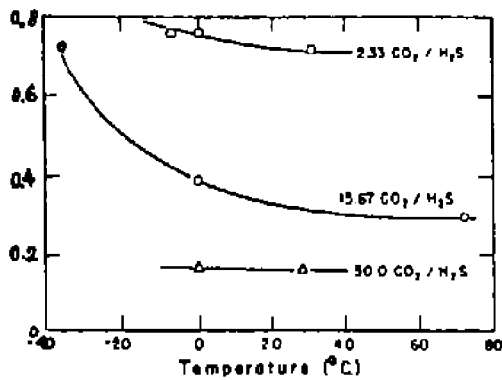


kg H<sub>2</sub>S Adsorbed/kg mol sieve



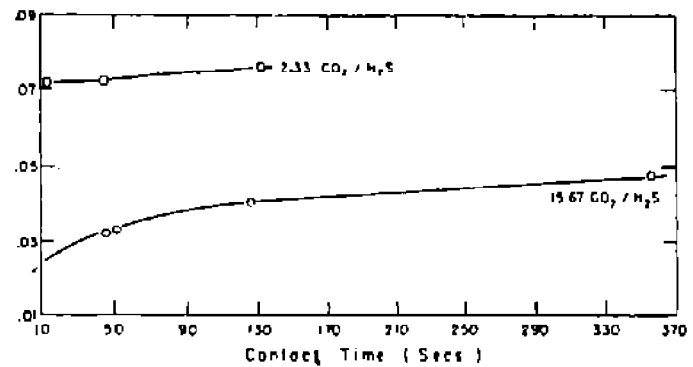
ADSORPTIVE CAPACITY OF MOLECULAR SIEVES AT 25°C  
 Figs. 1 & 2

kg H<sub>2</sub>S Adsorbed/kg mol sieve



EFFECT OF TEMPERATURE ON  
 ADSORPTIVE CAPACITY OF  
 MOLECULAR SIEVES AT  
 ATMOSPHERIC PRESSURE  
 Fig. 3

kg H<sub>2</sub>S Adsorbed/kg mol sieve



EFFECT OF CONTACT TIME ON  
 ADSORPTIVE CAPACITY OF MOLECULAR  
 SIEVES AT ATMOSPHERIC PRESSURE

Fig. 4

## 6.3 Process Design

### 6.3.1 Design philosophy

**6.3.1.1** The selected process should have given satisfactory service at process conditions and with gas compositions similar to that company's proposed treatment project.

**6.3.1.2** The Contractor shall pay particular attention to the heavy hydrocarbon analysis of the feed gas. If hydrocarbons condensed are absorbed in the treating solvent, severe process problems occur. Contractor's design shall incorporate features to remove or accommodate heavy hydrocarbons.

**6.3.1.3** Oxygen cause oxygation of certain treating solvents. Contractor should consider this in solvent selection if oxygen is in the feed gas.

**6.3.1.4** Solvent storage tanks shall be blanketed with sweet natural gas or inert gas. Vacuum system shall be avoided. If the solvent selected, is subject to oxygen degradation, provisions shall be made in design to prevent oxygen from entering the system.

**6.3.1.5** Solvent-acid gas loading shall be proportionated within accepted industry guidelines. However, for MEA, DEA, DGA and MDEA solvents the criteria given under Clauses 6.2.2.2, 6.2.2.3, 6.2.2.4 and 6.2.2.5 shall be considered respectively.

**6.3.1.6** For other proprietary solvents, the solvent to acid gas loading shall rigidly be based on the recommendations of the process Licensor(s).

**6.3.1.7** Solvent storage shall be provided with heating coils if freezing or high viscosity should prevent its normal transfer.

**6.3.1.8** Solvent filters shall be provided in accordance with accepted industrial practices.

**6.3.1.9** If the solvent selected, is capable of regeneration, complete regeneration facilities should be considered and designed.

**6.3.1.10** If proprietary processes are used, regeneration equipment shall be designed in strict accordance with Licensor's specification.

**6.3.1.11** Unless otherwise specified, acid gases shall be combusted in a flare or thermal oxidizer. Cold vents may be vented safely if the Company's approval is obtained.

**6.3.1.12** The Company shall specify the maximum allowable concentration of  $H_2S$ ,  $SO_2$  and such compounds at ground level. The guaranteed values should be specified in project specification.

**6.3.1.13** Due to considerable water vapor in treating plant, all essential determinations shall be considered in the process design, including but not limited to the followings:

- a)** The solvent stripping still shall be designed to prevent vacuum collapse in the event the tower is blocked in, when at a hot condition.
- b)** All equipment should be designed for potential vacuum collapse.
- c)** The acid gas disposal lines and facilities shall be designed so that water will not accumulate at the bottoms/lower ends.
- d)** Particular emphasis shall be given to lines in intermittent service such as drains, instruments, gage glasses, etc., to be freeze-protected.

**6.3.1.14** The following safety measures but not limited to, shall be taken into consideration in process design of the plant:

- a)** For toxic solvents (if selected), safety showers, eye-wash fountains, shall be provided at strategic locations.
- b)** All solvent and hydrocarbon tanks shall be diked to contain their contents in the event of a tank rupture.
- c)** An Emergency Shut-Down (ESD) system shall be provided if applicable. The ESD shall be automatically and manually actuated. The "ESD" system shall be designed to block the feed and products gas lines of (affected) train in the treating plant. The affected train shall be vented to the flare, at a rate not to damage the equipment. Gas flow to all fired heaters shall be stopped and all rotating equipment shall be stopped.
- d)** Plant pressure relief valves shall be separate from the ESD vents. ESD pushbuttons shall be provided at main gate, main control panel and at main entrance.
- e)** Hydrocarbon detectors, shall automatically trip the ESD system. They shall alarm at 20% of the lower explosive limit for methane and trip at 40%. They shall be provided at:
  - air intake to each control building;
  - analyzer house;
  - each compressor building;
  - in the gas contacting area;
  - above each cooling tower cell;
  - any other locations may deem necessary for safety requirements.
- f)** H<sub>2</sub>S detectors and alarms shall be provided if the gas to be treated contains more than 10 ppm<sub>v</sub> H<sub>2</sub>S. The detector shall have solid state sensors. The detector heads shall be located in proper places and the alarm system should have a two-level warning tone. One level shall be announced at an H<sub>2</sub>S concentration in ambient air of 10 ppm<sub>v</sub>. The other level shall be announced at 30 ppm<sub>v</sub> H<sub>2</sub>S in ambient air.
- g)** Equipment spacing within the process areas shall be in accordance with IPS-E-PR-190 "Layout and Spacing", unless not applicable for sour gas treating plants.

## 6.4 Material Selection for Sour Services

**6.4.1** The following requirements should be considered for vessels, drums, separator when they are used for sour services:

### **a) General**

Whenever the fluid to be handled in the process system contains:

- a.1)** sour gas of 0.34 kPa (abs) or greater partial pressure\* hydrogen sulfide (H<sub>2</sub>S), and if the system pressure is 450 kPa (abs) or greater.
- a.2)** Sour oil and multi-phase of 70 kPa (abs) or greater partial pressure H<sub>2</sub>S, whenever the system pressure is 1825 kPa (abs) or greater or whenever the gas phase contains over 15 percent H<sub>2</sub>S, at any system pressure.
- a.3)** Rich sweetening agents, such as amines, sulfinol, etc.

\* Partial pressure of H<sub>2</sub>S is determined by multiplying the mole fraction (mol. % +100) of H<sub>2</sub>S in the gas times the system pressure.



**b) Materials**

**b.1)** All carbon steel and low alloy steel material used in sour service shall comply with, but not be limited by, the requirements of NACE Standard MR-01-75. (Material Requirements-Sulfide Stress Cracking Resistant Material for Oilfield Equipment, latest Revision).

**b.2)** The use of special corrosion resistant materials, alloys or clad steels other than standard oilfield sour service materials, shall be subject to approval by the Company. Copper or copper alloys shall not be used.

**b.3)** The following Table 3 lists approved materials for use in sour services, based on ASTM specifications. The Vendor/Contractor may suggest alternate materials which are equal or better for consideration by the Company. These materials should also satisfy the physical and chemical properties requirements.

**TABLE 3 - ASTM SPECIFICATIONS**

PLATE	NO IMPACT TESTS REQUIRED	WHEN IMPACT TESTS REQUIRED
	A 516	A 516 (Normalized and impact tested to -45°C maximum)
Nozzles	A 106, Grade B	A 333, Grade 6
Flanges (Weld neck only)	A 105	A 350, Grade LF 2
Fittings	A 234, Grade WPB	A 420, Grade WPL 6
Bolts	A 193, Grade B7X/M	A 320, Grade L7M
Nuts	A 194, Grade 2HX/M	A 194, Grade 2M
Gaskets	TP-304 SS or TP-316 SS	TP-304 SS or TP-316 SS

**6.4.2 Piping for sour services**

**a)** Whenever the fluid handled in the process system contains:

**i)** Sour Gas of 0.34 kPa (abs) or greater partial pressure hydrogen sulphide (H<sub>2</sub>S), and if the system pressure is 450 kPa (abs) or greater.

**ii)** Sour oil and multiphase of 70 kPa (abs) or greater partial pressure H<sub>2</sub>S, whenever the system pressure is 1825 kPa (abs) or greater or, whenever the gas phase contains over 15 percent H<sub>2</sub>S at any system pressure.

**iii)** Rich sweetening agents, such as amines, sufinol, etc.

**b)** In general, the materials used in sour service shall be based on the following:

**i)** All carbon and low alloy steel material used in sour service shall comply with, but not limited, by the requirements of NACE Standard MR-01-75, latest revision.

**ii)** The use of special corrosion resistant materials, alloys or cold steels other than standard oil field sour service materials shall be subject to approval by the Company. Copper and copper alloys shall not be used.

**iii)** Stainless steels of the 400 series or other martensitic steels shall not be used unless specifically approved by the Company.

**iv)** All carbon and low alloy steel materials shall have the following properties:

- The ratio of percent manganese to percent carbon (%Mn / %C) shall be greater than or equal to 3.0.

- Percent carbon plus one-quarter of the percent manganese (%C + %Mn/4) shall be less than or equal to 0.55.

c) If specific materials are used in sour service piping components such as plates, pipes, fittings, flanges, valves, gaskets, etc., the applicable provisions of the following standards should be considered along with manufacturer's standard specification:

- IPS-M-PI-190
- ASTM-A 516, ASTM-A 106, Grade B, or ASTM-A 333 for plates and pipes respectively.
- ASTM-A 234 Grade WPB for fittings unless otherwise specified.
- ASTM-A 105 or ASTM-A 350, Grade LF2 for flanges unless otherwise specified.
- All valves shall meet the requirement of NACE Standard MR-01-75.
- All gaskets in sour service shall be TP 304 SS, TP 304 SS oval rings to ANSI B 16.20 shall be used as minimum.

## APPENDICES

### APPENDIX A

**TABLE A.1 - TYPICAL COMPARATIVE DATA OF MISCELLANEOUS PHYSICAL SOLVENTS**

PROCESS NAME	SELEXOL	FLUOR SOLVENT
Solvent name	Selexol	PC
FOB fact solvent cost \$/lb	1.32	0.74
Licenser	Norton	Fluor
Viscosity @ 25°C, cP.	5.8	3.0
Relative density (specific gravity) @ 25°C, kg/m <sup>3</sup>	1030	1195
Mol mass	280	102
Vapor pressure @ 25°C, mm Hg	$7.3 \times 10^{-4}$	$8.5 \times 10^{-2}$
Freezing point, °C	-28	-48
Boiling point, °C @ 760 mm Hg		240
Thermal conductivity W/m.k [Btu/Hr/Ft <sup>2</sup> /(°F/Ft)]	0.19 [0.11]	0.21 [0.12]
Maximum operating temp., °C	175	65
Specific Heat @ -4°C (25°F)	0.49	0.339
Water solubility @ 25°C	∞	94 g/L
Solvent solubility in water @ 25°C	∞	236 g/L
CO <sub>2</sub> solubility m <sup>3</sup> /L (Ft <sup>3</sup> /U.S. Gal) @ 25°C	0.0036 (0.485)	0.0034 (0.455)
Number of commercial plants	32	13
Bulk CO <sub>2</sub> removal		
Synthesis Gas	6	3
Natural Gas	6	10
Landfill Gas	3	0
Selective H <sub>2</sub> S removal		
Synthesis Gas	9	0
Natural Gas	8	0

## APPENDIX A

### TABLE A.2

PROCESS NAME	PURISOL	SEPASOLV MPE	ESTASOLVAN
Solvent name	NMP	Sepasolv	TBP
Solvent cost \$/lb FOB fact.	1.34		
Licensors	Lurgi	BASF	Uhde & IFP
Viscosity @ 25°C, cP.	1.65	—	2.9
Relative density (specific gravity) @ 25°C kg/m <sup>3</sup>	1027	—	973
Mol mass	99	320	266
Vapor pressure @ 25°C, mm Hg	$4.0 \times 10^{-1}$	$3.7 \times 10^{-4}$	$< 1.0 \times 10^{-2}$
Freezing point, °C	-24		-80
Boiling point, °C @ 760 mm Hg	202	320	(180 @ 30 mm Hg)
Thermal conductivity W/m.k [Btu/Hr/Ft <sup>2</sup> /(°F/Ft)]	0.16 [0.095]		
Maximum operating temp., °C	—	175	—
Specific heat @ -4°C (25°F)	0.40		
Water solubility @ 25°C			65 g/L
Solvent solubility in water @ 25°C			0.42 g/L
CO <sub>2</sub> solubility @ 25°C m <sup>3</sup> /L (Ft <sup>3</sup> /U.S. Gal)	0.0035 (0.477)	0.0034 (0.455)	0.0025 (0.329)
Number of commercial plants	5	4	0
Bulk CO <sub>2</sub> removal			
Synthesis Gas	2	0	
Natural Gas	1	0	
Landfill Gas	0	0	
Selective H <sub>2</sub> S removal			
Synthesis Gas	1	0	
Natural Gas	1	4	