ENGINEERING STANDARD

FOR

PROCESS DESIGN OF GAS TREATING UNITS

PART 1 - PROCESS DESIGN OF GAS SWEETENING UNITS

ORIGINAL EDITION

DECEMBER 1997

This standard specification is reviewed and updated by the relevant technical committee on Jan. 2005(1) and Dec. 2013(2). The approved modifications are included in the present issue of IPS.
FOREWORD

The Iranian Petroleum Standards (IPS) reflect the views of the Iranian Ministry of Petroleum and are intended for use in the oil and gas production facilities, oil refineries, chemical and petrochemical plants, gas handling and processing installations and other such facilities.

IPS is based on internationally acceptable standards and includes selections from the items stipulated in the referenced standards. They are also supplemented by additional requirements and/or modifications based on the experience acquired by the Iranian Petroleum Industry and the local market availability. The options which are not specified in the text of the standards are itemized in data sheet/s, so that, the user can select his appropriate preferences therein.

The IPS standards are therefore expected to be sufficiently flexible so that the users can adapt these standards to their requirements. However, they may not cover every requirement of each project. For such cases, an addendum to IPS Standard shall be prepared by the user which elaborates the particular requirements of the user. This addendum together with the relevant IPS shall form the job specification for the specific project or work.

The IPS is reviewed and up-dated approximately every five years. Each standards are subject to amendment or withdrawal, if required, thus the latest edition of IPS shall be applicable.

The users of IPS are therefore requested to send their views and comments, including any addendum prepared for particular cases to the following address. These comments and recommendations will be reviewed by the relevant technical committee and in case of approval will be incorporated in the next revision of the standard.

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GENERAL DEFINITIONS:
Throughout this Standard the following definitions shall apply.

COMPANY:
Refers to one of the related and/or affiliated companies of the Iranian Ministry of Petroleum such as National Iranian Oil Company, National Iranian Gas Company, National Petrochemical Company and National Iranian Oil Refinery And Distribution Company.

PURCHASER:
Means the “Company” where this standard is a part of direct purchaser order by the “Company”, and the “Contractor” where this Standard is a part of contract documents.

VENDOR AND SUPPLIER:
Refers to firm or person who will supply and/or fabricate the equipment or material.

CONTRACTOR:
Refers to the persons, firm or company whose tender has been accepted by the company.

EXECUTOR:
Executor is the party which carries out all or part of construction and/or commissioning for the project.

INSPECTOR:
The Inspector referred to in this Standard is a person/persons or a body appointed in writing by the company for the inspection of fabrication and installation work.

SHALL:
Is used where a provision is mandatory.

SHOULD:
Is used where a provision is advisory only.

WILL:
Is normally used in connection with the action by the “Company” rather than by a contractor, supplier or vendor.

MAY:
Is used where a provision is completely discretionary.
CONTENTS:

0. INTRODUCTION ................................................................. 1
1. SCOPE ............................................................................. 2
2. REFERENCES ................................................................. 2
3. DEFINITIONS AND TERMINOLOGY ................................... 3
4. SYMBOLS AND ABBREVIATIONS ...................................... 5
5. UNITS .............................................................................. 5

PART I .................................................................................. 6
6. GAS SWEETENING UNITS .................................................. 6
   6.1 General ................................................................. 6
   6.2 Process Selection .................................................... 6
      6.2.1 General .......................................................... 6
      6.2.2 Chemical reaction processes ............................ 7
      6.2.3 Combined physical/chemical purification processes 15
      6.2.4 Carbonate process .......................................... 16
      6.2.5 Catacarb process ............................................. 17
      6.2.6 Physical absorption methods ............................ 18
   6.3 Process Design ......................................................... 22
      6.3.1 Design philosophy .......................................... 22
   6.4 Material Selection for Sour Services ......................... 23

APPENDICES:

APPENDIX A GUIDELINES FOR SELECTION OF SOLVENT-BASED PROCESS .......... 26
APPENDIX B TYPICAL AMINE PLANT SHOWING PRINCIPAL AREAS OF CORROSION AND SUGGESTED MATERIALS OF CONSTRUCTIONS .............................. 29
APPENDIX C TYPICAL COMPARATIVE DATA OF MISCELLANEOUS ......................... 30
0. INTRODUCTION

Design of Gas Treating Process is broad and contains subjects of paramount importance. Therefore this engineering IPS standards specification is prepared to cover the subject.

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 treating 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 only the part I has been discussed in this Standard:

Part I: “Process Design of Gas Sweetening Units”

Part II: “Process Design of Gas Dehydration Units”


Note 1:
This standard specification is reviewed and updated by the relevant technical committee on Jan. 2005. The approved modifications by T.C. were sent to IPS users as amendment No. 1 by circular No. 253 on Jan. 2005. These modifications are included in the present issue of IPS.

Note 2:
This standard specification is reviewed and updated by the relevant technical committee on Dec. 2013. The approved modifications by T.C. were sent to IPS users as amendment No. 2 by circular No. 401 on Dec. 2013. These modifications are included in the present issue of IPS.

2. REFERENCES

Throughout this Standard the following dated and undated standards/codes are referred to. These referenced documents shall, to the extent specified herein, form a part of this standard. For dated references, the edition cited applies. The applicability of changes in dated references that occur after the cited date shall be mutually agreed upon by the Company and the Vendor. For undated references, the latest edition of the referenced documents (including any supplements and amendments) applies.

IPS (IRANIAN PETROLEUM STANDARDS)

IPS-E-GN-100 “Engineering Standard for Units”
IPS-M-PI-190 “Material and Equipment Standard for Line Pipe”
IPS-E-PR-190 “Engineering Standard for Layout and Spacing”

ASTM (AMERICAN SOCIETY FOR TESTING AND MATERIALS)

ASTM A516/ A516 M-2012 “Standard Specification for Pressure Vessel Plates, Carbon Steel, for Moderate- and Lower-Temperature Service”
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 CO₂, H₂S, COS, RSH and SO₂. Most common in natural gas are CO₂, H₂S and 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 Amine (Alkanolamine)
Any of several liquid compounds containing amino nitrogen generally used in water solution to remove, by reversible chemical reaction, hydrogen sulfide and/or carbon dioxide from gas and liquid hydrocarbon streams.

3.6 Antifoam
Antifoam is a substance, usually a silicone or long-chain alcohol, added to the treating system to reduce the tendency to foam.
3.7 Degradation Products
Degradation products are impurities in a treating solution which are formed from both reversible and irreversible side reactions.

3.8 Liquefied Petroleum Gas (LPG)
Predominately propane or butane, either separately or in mixtures, which is maintained in a liquid state under pressure within the confining vessel.

3.9 Mercaptan
Mercaptan is a hydrocarbon group (usually a methane, ethane, or propane) with a sulfur group (–SH) substituted on a terminal carbon atom.

3.10 Natural Gas
Natural gas is a gaseous form of petroleum obtained from underground sources and consisting of a complex mixture of hydrocarbons, primarily ethane, but generally also including ethane, propane and higher hydrocarbons in much smaller amounts. It generally also includes some inert gases, such as nitrogen and carbon dioxide, plus minor amounts of trace constituents.

3.11 Normal cubic meters (Nm³)
Refers to capacity at Normal Conditions (101.325 kPa & 0°C) and relative humidity of 0 percent.

3.12 Pressure Relief Valve
Valve designed to open and relieve excess pressure and to reclose and prevent the further flow of fluid after normal conditions have been restored.

3.13 Reclaimer
A kind of heat exchanger which is used to remove degradation products from the solution and also aids in the removal of heat stable salts, suspended solids, acids and iron compounds.

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

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

3.16 Sour Gas
Gas containing undesirable quantities of hydrogen sulfide, mercaptans, and/or carbon dioxide. It also is used to refer to the feed stream to a sweetening unit.

3.17 Standard Cubic Meters (SCM)
Refers to capacity at standard conditions (101,325 kPa & 15.5°C)
3.18 Steric Hinderance
Steric hindrance is chemically attaching a bulky molecule (such as benzene) to the hydrocarbon chain of an amine to inhibit CO₂ reacting to form a carbonate.

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

3.20 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
CO₂ Carbon Dioxide.
COS Carbonyl Sulfide.
CS₂ Carbon Disulfide.
DEA di-Ethanolamine.
DETA di-Ethylamine.
DIPA di-Isopropanolamine.
DGA di-Glycolamine.
ESD Emergency Shut-Down.
FOB Free On Board.
GPSA Gas Processors Suppliers Association.
HCN Hydrogen Cyanide.
H₂S Hydrogen Sulfide.
LPG Liquefied Petroleum Gas.
MEA mono-Ethanolamine.
MDEA Methyl di-Ethanolamine.
max. maximum.
min. minimum.
NACE National Association of Corrosion Engineers.
ppm_w parts per million by weight, in (mg/kg).
ppm_v parts per million by volume, in (µm³/m³).
SS Stainless Steel.
TLV Threshold Limit Value.

5. UNITS
This Standard is based on International System of Units (SI), as per IPS-E-GN-100 except where otherwise specified.
6. GAS SWEETENING UNITS

6.1 General

Acid gas constituents present in most natural gas streams are mainly hydrogen sulfide (H₂S) and carbon dioxide (CO₂). 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 shall meet the desired 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 H₂S, 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, cost, 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, it's 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 H₂S and CO₂ removal or selective H₂S absorption is desired.

A guideline for selection of solvent-based process is shown in appendix A. For more information refer to Fundamental of Natural Gas Processing by Arthur J. Kidnay William R. Parrish, Pages 96-97/Gas Conditioning and processing by Maddox Vol. 4, Pages 8-10.

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:

- The environmental constrains, including air pollution regulations and disposal of by products considered hazardous chemicals;
- Type and concentration of impurities in sour gas;
- Specification of treated gas (sweet 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, operating and royalty cost for process;
- Liquid product specifications (where applicable).
6.2.1.5 An accurate analysis of the sour gas stream showing all of the impurities including COS, CS₂ 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 downstream 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 selection 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.

6.2.1.8 The feasibility and desirability of sulfur recovery can also place considerable limitation on the selection of sweetening process. Removal of H₂S to produce contractually sweetened gas is difficult 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 H₂S and/or CO₂ 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 H₂S and CO₂ from natural gas streams due to their reactivity and availability at low cost. Typical gas sweetening by chemical reaction is shown in appendix A.

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 sweet gas. Physical properties for some of the gas treating chemicals are shown in Table 1. The alkanolamine 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

<table>
<thead>
<tr>
<th>Physical Property</th>
<th>Propylene Carbonate</th>
<th>Methyl Diethanolamine</th>
<th>Sulfolane</th>
<th>Methanol</th>
<th>10% Sodium Hydroxide</th>
</tr>
</thead>
<tbody>
<tr>
<td>Formula</td>
<td>C₃H₄CO₂</td>
<td>(HOC₃H₄NCH₂)</td>
<td>C₃H₇SO₂</td>
<td>CH₃OH</td>
<td>C₃H₇OH</td>
</tr>
<tr>
<td>Molecular Wt.</td>
<td>102.09</td>
<td>110.16</td>
<td>120.17</td>
<td>32.04</td>
<td>19.05</td>
</tr>
<tr>
<td>Boiling point @ 760 mm Hg, °C</td>
<td>241</td>
<td>247</td>
<td>285</td>
<td>64.5</td>
<td>102.8</td>
</tr>
<tr>
<td>Freezing point, °C</td>
<td>-49.2</td>
<td>-23</td>
<td>27.6</td>
<td>-97.7</td>
<td>-10</td>
</tr>
<tr>
<td>Critical constants</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure, kPa (abs)</td>
<td>-</td>
<td>-</td>
<td>-52.09</td>
<td>7066</td>
<td></td>
</tr>
<tr>
<td>Temperature, °C</td>
<td>-</td>
<td>-</td>
<td>545</td>
<td>240</td>
<td></td>
</tr>
<tr>
<td>Density @ 20°C, gm/cc.</td>
<td>1.2037</td>
<td>1.040</td>
<td>1.288</td>
<td>1.110</td>
<td></td>
</tr>
<tr>
<td>Weight, kg/m³</td>
<td>1040</td>
<td>1273 @ 30°C/30°C</td>
<td>792</td>
<td>1109</td>
<td></td>
</tr>
<tr>
<td>Relative density 20°C/20°C</td>
<td>1.203</td>
<td>1.0415</td>
<td>1.268</td>
<td>1.110</td>
<td></td>
</tr>
<tr>
<td>Specific heat @ 15.6°C, kJ/kg °C</td>
<td>1.40</td>
<td>2.24</td>
<td>1.47 @ 30°C</td>
<td>2.47 @ 5-10°C</td>
<td>3.76</td>
</tr>
<tr>
<td>Thermal conductivity</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>W/m °C @ 20°C</td>
<td>0.21 @ 10°C</td>
<td>0.275</td>
<td>0.197 @ 38°C</td>
<td>0.215</td>
<td></td>
</tr>
<tr>
<td>Latent heat of vaporization, kJ/kg</td>
<td>484 @ 760 mm Hg</td>
<td>476</td>
<td>525 @ 100°C</td>
<td>1103 @ 760 mm Hg</td>
<td></td>
</tr>
<tr>
<td>Heat of reaction, kJ/kg of Acid Gas</td>
<td>H₂S</td>
<td>-</td>
<td>-52.09</td>
<td>7066</td>
<td></td>
</tr>
<tr>
<td>CO₂</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Viscosity, m²/s</td>
<td>1.57 @ 38°C</td>
<td>1.3 x 10⁻⁶ m²/s @ 10°C</td>
<td>10.5 @ 30°C</td>
<td>0.6 @ 20°C</td>
<td>1.83 @ 20°C</td>
</tr>
<tr>
<td></td>
<td>19.4 x 10⁻⁶ m²/s @ -40°C</td>
<td>6.8 x 10⁻⁶ m²/s @ 38°C</td>
<td>6.1 @ 50°C</td>
<td>0.97 @ 50°C</td>
<td></td>
</tr>
<tr>
<td></td>
<td>1.7 x 10⁻⁶ m²/s @ 38°C</td>
<td>0.28 x 10⁻⁶ m²/s @ 100°C</td>
<td>2.5 @ 100°C</td>
<td>0.40 @ 100°C</td>
<td></td>
</tr>
<tr>
<td></td>
<td>0.82 x 10⁻⁶ m²/s @ 99°C</td>
<td>1.4 @ 150°C</td>
<td>0.97 @ 200°C</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Refractive index, N₀ @ 20°C</td>
<td>1.4209</td>
<td>1.489</td>
<td>1.481 @ 30°C</td>
<td>1.3286</td>
<td></td>
</tr>
<tr>
<td>Flash point, °C</td>
<td>132.2</td>
<td>129.4</td>
<td>177</td>
<td>14</td>
<td></td>
</tr>
</tbody>
</table>

* COC: Cleveland Open Cap is a test method
6.2.2.2 Mono-Ethanolamine (MEA)

6.2.2.2.1 MEA should commonly be used as a 15 to 25% 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, CS₂, SO₂ and SO₃ can partially deactivate MEA, which may essentially require to be recovered with a reclaimer.

6.2.2.2.3 Since MEA is a primary amine, it has a high pH. This enables MEA solutions to produce gas containing less than 6 mg/Sm³ (¼ grains H₂S per 100 scu.ft) of acid gas at very low H₂S partial pressures. The reaction of MEA with Oxygen forms corrosive thiosulfate.

6.2.2.2.4 The heat of reaction for CO₂ in MEA is about 1930 kJ/kg of CO₂ (460 kcal/kg of CO₂). 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 H₂S/Sm³ 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 H₂S/Sm³ 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 0.35 to 0.8 moles acid gas/mole of amine 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 CS₂ 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 H₂S and CO₂. As a result, for some low pressure gas streams, DEA can not produce Pipeline Specification gas. However, certain design arrangement such as split flow may be considered to fulfill the specified requirement.

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 H₂S and will permit a significant fraction of CO₂ to remain in the product gas.

6.2.2.3.6 The heat of reaction for DEA and CO₂ is 1515 kJ/kg of CO₂ (360 kcal/kg of CO₂) 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 50 to 70 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 CO₂ over H₂S. It also has a higher pH than MEA and thus can easily achieve 6 mg H₂S/Sm³ gas (0.25 grains per 100 scu.ft) except in some cases where large amounts of CO₂ are present relative to H₂S.

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. Some properties of DGA in comparison
with other amines are presented in Table 2.

*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 40 to 50 mass percent range. Due to considerably reduced corrosion problems, acid gas loadings as high as 0.4 to 0.55 mole acid gas/mole amine are practical in carbon steel equipment.

6.2.2.5.2 Since MDEA is a tertiary amine, it has less affinity for H₂S and CO₂ 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 H₂S in the presence of CO₂.

**TABLE 2 - COMPARATIVE DATA TABLE**

<table>
<thead>
<tr>
<th>AMINE</th>
<th>MEA</th>
<th>DEA</th>
<th>DGA</th>
<th>MDEA</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solution strength, mass %</td>
<td>15-25</td>
<td>25-35</td>
<td>50-70</td>
<td>40-50</td>
</tr>
<tr>
<td>Acid gas loading mole acid gas/mole amine</td>
<td>0.3-0.4</td>
<td>0.35-0.8</td>
<td>0.3-0.4</td>
<td>0.4-0.55</td>
</tr>
<tr>
<td>*AHR for H₂S kJ/kg (kcal/kg)</td>
<td>1281 (306)</td>
<td>1189 (284)</td>
<td>1570 (375)</td>
<td>1214 (290)</td>
</tr>
<tr>
<td>*AHR for CO₂ kJ/kg (kcal/kg)</td>
<td>1922 (459)</td>
<td>1520 (363)</td>
<td>1729 (413)</td>
<td>1398 (334)</td>
</tr>
<tr>
<td>Ability to preferentially absorb H₂S</td>
<td>No</td>
<td>Under some conditions</td>
<td>No</td>
<td>Under most conditions</td>
</tr>
</tbody>
</table>

*Approximate Heat of Reaction: The heat of reaction varies with acid gas loading and solution concentration.

6.2.2.6 Mixed Amines/Formulated Solvent

6.2.2.6.1 Mixed Amines

The selectivity of MDEA can be reduced by addition of MEA, DEA, or proprietary additives. Thus, it can be tailored to meet the desired amount of CO₂ slippage and still have lower energy requirements than do primary and secondary amines. (Refer to: Fundamentals of Natural Gas Processing, Ed.2006, sec. 5.2.1.7)

6.2.2.6.2 Formulated Solvent

Formulated Solvents is the name given to a new family of amine-based solvents. Their popularity is primarily due to equipment size reduction and energy savings over most of the other amines. All the advantages of MDEA are valid for the Formulated Solvents, usually to a greater degree. Some formulations are capable of slipping larger portions of inlet CO₂ (than MDEA) to the outlet gas and at the same time removing H₂S to less than 4 ppmv. (Refer to: GPSA Data Book “Gas Processors
6.2.2.6.3 Some advantage of Mixed/Formulated Amines are:

For New Plants:
- reduced corrosion for formulated solvents
- reduced circulation rate
- lower energy requirements
- smaller equipment due to reduced circulation rates

For Existing Plants:
- increase in capacity, i.e., gas through-put or higher inlet acid gas composition
- reduced corrosion for formulated solvents
- lower energy requirements and reduced circulation rates


6.2.2.7 General considerations

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:

6.2.2.7.1 Corrosion

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

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

c. Lowest possible pressure on stripping columns and reboilers should be considered to avoid severe corrosion of reboiler tubes.

d. 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.

e. Most of the equipment and piping in an alkanolamine plant is constructed of carbon steel. It is common practice to construct certain sections of an amine plant with stainless steel or other alloy. Selective use of corrosion resistant alloys permits operation at higher amine concentrations and acid gas loadings, allows better stripping of the lean amine solution and improved treating, reduces corrosion in susceptible areas, and improves process economics. Locations marked in bold on Appendix B Fig.1 show where carbon steel is typically replaced by stainless steel.

Typical amine plant showing principal areas of corrosion and suggested materials of constructions shown in appendix B Fig.1.

6.2.2.7.2 Heat-Stable Salts

Carbon dioxide and hydrogen sulfide are weak enough acids that their reactions with amines are thermally reversible. Acids that are sufficiently strong that their reactions with amines are not
thermally reversible are called heat-stable acids, and products of their reactions with amines are heat-stable salts. If heat-stable acids enter an amine unit or are generated in the amine solution by reaction with trace amounts of oxygen or by thermal degradation of the amine, heat-stable salts can accumulate in the solution. Heat-stable salts reduce the acid gas removal capacity and increase the corrosivity of amine solutions by tying up a portion of the amine. Corrosion due to heat stable salts can be controlled by amine reclaiming and/or the addition of soda ash or caustic soda to neutralize the acids involved.

6.2.2.7.3 Foaming  

a. A sudden increase in differential pressure across a contactor often indicates severe foaming. When foaming occurs, there is poor contact between the gas and the chemical solution. The result is reduced treating capacity and sweetening efficiency, possibly to the point that outlet specification cannot be met. Some reasons for foaming are:

- Suspended solids
- Organic acids
- Corrosion inhibitors
- Condensed hydrocarbons
- Soap-based valve greases
- Makeup water impurities
- Degradation products
- Lube Oil

b. Foaming problems can usually be traced to plant operational problems. Contaminants from upstream operations can be minimized through adequate inlet separation. Condensation of hydrocarbons in the contactor can usually be avoided by maintaining the lean solution temperature at least 5°C above the hydrocarbon dew point temperature of the outlet gas. Temporary upsets can be controlled by the addition of antifoam chemicals. These antifoams are usually of the silicone or long-chain alcohol type.

6.2.2.7.4 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.

6.2.2.7.5 Absorber  

a. Usually the absorber is a tray column, although packed columns are also used. In either case, the objective is to provide intimate contact between the gas and the amine solvent so that the H₂S and CO₂ molecules can transfer from the gas phase to the solvent liquid phase.

b. The following criteria are minimum requirements to be consider in the absorber columns:

- Design the absorber with adequate vapor cross section area to prevent amine mechanical entrainment and loss.
- The inlet temperature of the absorber feed should be maintained as low as practical.
- For hydrocarbon gas, the lean solution temperature must be maintained at least 5°F above the gas feed temperature at the absorber inlet.
• For gas treating, the absorber should be designed to operate at the highest practical pressure, within the constraints imposed by needs of the upstream and downstream units.

6.2.2.7.6 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.

6.2.2.7.7 Filtration

a. 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.

b. 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.

6.2.2.7.8 Amine reclaiming

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

1. 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.

2. 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.

3. 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.

4. The tubes should be widely spaced for easy cleaning.

5. Make certain the steam supply is not superheated. Usually, to make saturated steam, the BFW is injected to the superheated steam at the inlet of the reclaimer tube.

6. A recorder should be used to monitor the reclaimer temperature throughout the cycle.

7. 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. Temperature indicator should be provided on reclaimer outlet line.

8. Sufficient vapor space should be allowed above the liquid layer in the reclaimer to prevent liquid carryover in the overhead vapor line.

9. 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.
6.2.2.7.9 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. In case of using shell and tube heat exchanger, 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 offset the increased investment. Flowing amine solutions should not impinge directly on vessel surfaces; impingement baffles should be utilized in the exchangers.

6.2.2.7.10 Reboiler of regenerator

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

a. 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.

b. 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 CO2. However increased corrosion in the reboiler, exchanger and stripper has been reported with the higher reactivator operating pressure. Therefore, it may be necessary to operate at a minimum reactivator pressure, with the lower limit set usually either by the pressure required to supply the acid gas to a sulfur plant or by the pressure required to force the amine out of the regenerator through the heat exchanger to a surge tank or pump suction.

c. The maximum allowable kettle temperature recommended for regeneration of the solution depends on the type of Amine used to prevent amine degradation.

d. 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.

e. 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.

f. To provide good circulation of the amine solution around the tubes and to reduce fouling causes by a sludge accumulation, the tube bundle should be placed on a slide about 15 cm above the bottom of the reboiler shell.

g. 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.

h. The tube bundle should be supported to prevent tube vibration. Teflon or other protective inserts will prevent tube cutting.

i. 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.

j. 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.

k. 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.

l. 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.

m. 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.

6.2.2.7.11 Amine solution selection

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

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 H2S removal from streams containing CO2 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 H2S.

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 CO2 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 CO2 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$_2$S and COS are absorbed together with CO$_2$ 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 - DIPA.
2) Di-Ethylamine - DETA

6.2.3.3 These new DIPA 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$_2$ and H$_2$S loadings.
c) High selectivity of the solvent between sulfur components and CO$_2$, 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$_2$, along with very low residual sulfur contents in the clean gas. Even with very low sulfur and high CO$_2$ contents in feed gas, it is unproblematic to produce a high sulfur off-gas containing more than 25% H$_2$S, which is suitable for Claus plants.

6.2.3.5 The CO$_2$ produced by alkylamine process is free from COS and shall meet today’s stringent environmental pollution requirements. CO$_2$ 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$_2$S, COS and CO$_2$ as well as trace contaminants, such as HCN, NH$_3$, 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$_2$ 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$_2$ from a number of gas mixtures. It has been used for sweetening natural gases containing both CO$_2$ and H$_2$S.

6.2.4.2 Hot Potassium Carbonate Process is not suitable for sweetening gas mixtures containing little or not CO$_2$, as the potassium bisulfide should be very difficult to regenerate if CO$_2$ 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 H₂S 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 Successful corrosion inhibitors that so far reported includes arsenic and vanadium salts as well as dichromates.

6.2.4.4.2 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.

6.2.4.4.3 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.

6.2.4.4.4 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 H₂S is present in the gas, chromates cannot be used. Both H₂S 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 H₂S is present, formation of the potassium bisulfide inhibits corrosion to a significant extent. In cases where corrosion is a problem with H₂S present, the commercially available film-forming fatty amines are usually effective as corrosion inhibitors.

6.2.4.4.5 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.

6.2.4.4.6 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 nontoxic 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, CS₂ 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-Sec. 21-P: 21-13.

b) The heavy hydrocarbon concentration in the feed gas is low.

c) Bulk removal of the acid gas is desired.

d) Selective removal of H₂S is desired.

In general, physical solvents are capable of removing COS, CS₂ and mercaptans.

6.2.6.2 Solvent regeneration

6.2.6.2.1 The solvents are regenerated by:

a) Multi-stage flashing 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) 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 H₂S or mercaptans. The solids processes tend to be highly selective and do not normally remove significant quantities of CO₂. Consequently, the H₂S 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 Sm³ [1000 grains H₂S per 100 Scf] of gas.

6.2.7.2.3 This process consists of wood chips impregnated with varying amounts of hydrate iron oxide (Fe₂O₃). This reacts with the H₂S to form Fe₂S₃, 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.

b) 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 alumino silicates can be used for selective removal of H₂S 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 H₂S gas approaches about 0.15 kg (pounds) of H₂S 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 H₂S is in a mixture rather than as a pure component.

Fig. 3 shows that the adsorptive capacity of molecular sieves for H₂S increases as temperature is lowered. Also shown is the effect of CO₂/H₂S ratio. As the percentage H₂S in the mixture increases the adsorptive capacity of the molecular sieve for H₂S 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.

![Graphs of Adsorptive Capacity of Molecular Sieves](image-url)
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 Particular attention shall be paid to the heavy hydrocarbons analysis of the feed gas. If
condensed hydrocarbons are absorbed in the treating solvent, severe process problems occur.
Provisions shall be made in design to remove or accommodate heavy hydrocarbons.

6.3.1.3 This should be considered insolvent 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 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 should 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, can be regenerated, complete regeneration facilities shall 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 Acid gases may be combusted in a flare or thermal oxidizer if compatible with the
environmental regulations upon Company’s approval.

6.3.1.12 The Company shall specify the maximum allowable concentration of H₂S, SO₂ 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 shall be taken into consideration in process design of the plant as a minimum:

   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 separated from the ESD vents. ESD push buttons shall be provided at main gate, main control panel and at main entrance.

   e) Unless otherwise specified by company, hydrocarbon detectors shall activate the ESD system, automatically. 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) H2S detectors and alarms shall be provided if the gas to be treated contains more than 10 ppmv H2S. 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 H2S concentration in ambient air of 10 ppmv. The other level shall be announced at 30 ppmv H2S 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 and separators when they are used for sour services:

   a) General

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

   a.1) ≥ 50 ppmw total sulfide content in the aqueous phase; or
a.2) ≥1 ppmw total sulfide content in the aqueous phase and pH < 4; or

a.3) ≥1 ppmw total sulfide content and > 20 ppmw free cyanide in the aqueous phase, and pH > 7.6; or

a.4) > 0.3 kPa absolute (0.05 psia) partial pressure H2S in the gas phase associated with the aqueous phase of a process.

* Partial pressure of H2S is determined by multiplying the mole fraction (mol. % +100) of H2S in the gas times the system pressure.

Note:
The high-PH sour environments differentiate refinery sour service from the oil and gas production sour environments covered by NACE MR0175/ISO 15156, because many wet sour streams in production also contain carbon dioxide and hence exhibit a lower PH. Another major difference is that chloride ion concentrations tend to be significantly lower in refinery sour services than in oil production sour services.

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 0103. (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. (to see more detail refer to IPS-E-TP-760, Sec. 5.3/5.4; and IPS-E-TP-740, App., B.1.11/12).

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

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 (H2S), and if the system pressure is 450 kPa (abs) or greater.

ii) Sour oil and multiphase of 70 kPa (abs) or greater partial pressure H2S, whenever the system pressure is 1825 kPa (abs) or greater or, whenever the gas phase contains over 15 percent H2S at any system pressure.
iii) Rich sweetening agents, such as amines.

b) In general, the materials used in sour service shall be based on the following:
   i) All carbon steel and low alloy steel material used in sour service shall comply with, but not
      limited, by the requirements of NACE Standard MR 0103, 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 steel 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
       - Percent carbon plus one-quarter of the percent manganese (%C + %Mn/4) shall be less
       equal to 3.0.
       than or equal to 0.55.

c) For materials used in sour service piping components such as plates, pipes, fittings, flanges,
   valves, gaskets, etc., the applicable provisions of the following standards shall be considered
   along with manufacturer’s standard specification:
   • IPS-M-PI-190
   • API 5L
   • 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 0103.
   • All gaskets in sour service shall be TP 304 SS, TP 304 SS oval rings to ASME B 16.20 shall
     be used as minimum.

6.4.3 A typical material selection diagram is presented in Appendix B.
APPENDICES

APPENDIX A

GUIDELINES FOR SELECTION OF SOLVENT-BASED PROCESS

Fig. A.1

PRIMARY SOURCES OF ENERGY IN THE WORLD IN 2003. TOTAL ENERGY USED WAS 405 QUADRILLION BTU (ENERGY INFORMATION ADMINISTRATION, 2005b).

Fig. A.2

MAJOR PROVEN NATURAL GAS RESERVES BY COUNTRY. TOTAL PROVEN RESERVES ESTIMATED TO BE 6,040 Tcf (ENERGY INFORMATION ADMINISTRATION, 2005c).
UNITED STATES ENERGY CONSUMPTION BY FUEL. (ADAPTED FROM ENERGY INFORMATION ADMINISTRATION, 2005a.)

Fig. A.3

NATURAL GAS SUPPLY AND DISPOSITION IN THE UNITED STATES IN 2003. VALUES SHOWN ARE IN TCF. (ADAPTED FROM ENERGY INFORMATION ADMINISTRATION, 2005d.)

Fig. A.4
TYPICAL GAS SWEETENING BY CHEMICAL REACTION

Fig. A.5
APPENDIX B
TYPICAL AMINE PLANT SHOWING PRINCIPAL AREAS OF CORROSION AND SUGGESTED MATERIALS OF CONSTRUCTIONS

FLOW DIAGRAM OF A TYPICAL AMINE PLANT SHOWING PRINCIPAL AREAS OF CORROSION AND SUGGESTED MATERIALS OF CONSTRUCTION

Fig. B.1
# APPENDIX C

## TYPICAL COMPARATIVE DATA OF MISCELLANEOUS PHYSICAL SOLVENTS PROCESSES

### TABLE C.1 - TYPICAL COMPARATIVE DATA OF MISCELLANEOUS PROCESS NAME

<table>
<thead>
<tr>
<th>PROCESS NAME</th>
<th>SELEXOL</th>
<th>FLUOR SOLVENT</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solvent name</td>
<td>Selexol</td>
<td>PC</td>
</tr>
<tr>
<td>FOB fact solvent cost $/lb</td>
<td>*</td>
<td>*</td>
</tr>
<tr>
<td>Licensor</td>
<td>Norton</td>
<td>Fluor</td>
</tr>
<tr>
<td>Viscosity @ 25°C, cP.</td>
<td>5.8</td>
<td>3.0</td>
</tr>
<tr>
<td>Relative density (specific gravity) @ 25°C, kg/m³</td>
<td>1030</td>
<td>1195</td>
</tr>
<tr>
<td>Mol mass</td>
<td>280</td>
<td>102</td>
</tr>
<tr>
<td>Vapor pressure @ 25°C, mm Hg</td>
<td>7.3 x 10^-4</td>
<td>8.5 x 10^-2</td>
</tr>
<tr>
<td>Freezing point, °C</td>
<td>-28</td>
<td>-48</td>
</tr>
<tr>
<td>Boiling point, °C @ 760 mm Hg</td>
<td></td>
<td>240</td>
</tr>
<tr>
<td>Thermal conductivity W/m.k [Btu/Hr/Ft²/(°F/Ft)]</td>
<td>0.19 [0.11]</td>
<td>0.21 [0.12]</td>
</tr>
<tr>
<td>Maximum operating temp., °C</td>
<td>175</td>
<td>65</td>
</tr>
<tr>
<td>Specific Heat @ -4°C (25°F)</td>
<td>0.49</td>
<td>0.339</td>
</tr>
<tr>
<td>Water solubility @ 25°C</td>
<td>∞</td>
<td>94 g/L</td>
</tr>
<tr>
<td>Solvent solubility in water @ 25°C</td>
<td>∞</td>
<td>236 g/L</td>
</tr>
<tr>
<td>CO₂ solubility m³/L (Ft³/U.S. Gal)</td>
<td>0.0036 (0.485)</td>
<td>0.0034 (0.455)</td>
</tr>
<tr>
<td>Number of commercial plants</td>
<td>32</td>
<td>13</td>
</tr>
<tr>
<td>Bulk CO₂ removal Synthesis Gas</td>
<td>6</td>
<td>3</td>
</tr>
<tr>
<td>Natural Gas</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>Landfill Gas</td>
<td>3</td>
<td>0</td>
</tr>
<tr>
<td>Selective H₂S removal Synthesis Gas</td>
<td>9</td>
<td>0</td>
</tr>
<tr>
<td>Natural Gas</td>
<td>8</td>
<td>0</td>
</tr>
</tbody>
</table>

**Note:** (*) To be completed at the time of comparison.
## TABLE C.2

<table>
<thead>
<tr>
<th>PROCESS NAME</th>
<th>PURISOL</th>
<th>SEPASOLV MPE</th>
<th>ESTASOLVAN</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solvent name</td>
<td>NMP</td>
<td>Sepasolv</td>
<td>TBP</td>
</tr>
<tr>
<td>Solvent cost $/lb FOB fact.</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Licensor</td>
<td>Lurgi</td>
<td>BASF</td>
<td>Uhde &amp; IFP</td>
</tr>
<tr>
<td>Viscosity @ 25°C, cP.</td>
<td>1.65</td>
<td>---</td>
<td>2.9</td>
</tr>
<tr>
<td>Relative density (specific gravity) @ 25°C kg/m³</td>
<td>1.027</td>
<td>---</td>
<td>0.973</td>
</tr>
<tr>
<td>Mol mass</td>
<td>99</td>
<td>320</td>
<td>266</td>
</tr>
<tr>
<td>Vapor pressure @ 25°C, mm Hg</td>
<td>4.0 × 10⁻¹</td>
<td>3.7 × 10⁻¹</td>
<td>&lt; 1.0 × 10⁻¹</td>
</tr>
<tr>
<td>Freezing point, °C</td>
<td>-24</td>
<td>-80</td>
<td></td>
</tr>
<tr>
<td>Boiling point, °C @ 760 mm Hg</td>
<td>202</td>
<td>320</td>
<td>(180 @ 30 mm Hg)</td>
</tr>
<tr>
<td>Thermal conductivity W/m.k [Btu/Hr/Ft²/(°F/Ft)]</td>
<td>0.16</td>
<td>320</td>
<td>(0.095)</td>
</tr>
<tr>
<td>Maximum operating temp., °C</td>
<td>---</td>
<td>175</td>
<td>---</td>
</tr>
<tr>
<td>Specific heat @ -4°C (25°F)</td>
<td>0.40</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water solubility @ 25°C</td>
<td></td>
<td>65 g/L</td>
<td></td>
</tr>
<tr>
<td>Solvent solubility in water @ 25°C</td>
<td></td>
<td>0.42 g/L</td>
<td></td>
</tr>
<tr>
<td>CO₂ solubility m³/L (Ft³/U.S. Gal) @ 25°C</td>
<td>0.0035</td>
<td>0.0034</td>
<td>0.0025</td>
</tr>
<tr>
<td>Number of commercial plants</td>
<td>5</td>
<td>4</td>
<td>0</td>
</tr>
<tr>
<td>Bulk CO₂ removal</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Synthesis Gas</td>
<td>2</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Natural Gas</td>
<td>1</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Landfill Gas</td>
<td>0</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Selective H₂S removal</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Synthesis Gas</td>
<td>1</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Natural Gas</td>
<td>1</td>
<td>4</td>
<td></td>
</tr>
</tbody>
</table>

Note: (*) To be completed at the time of comparison.