FUNDAMENTALS OF GAS TREATING

by

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INTRODUCTION

In this paper it is our intention to review the fundamentals of gas treating by liquid and solid processes. In the first part we will describe the Ethanolamine Process for treating gas to remove acid gas constituents, proprietary liquid processes including Diglycolamine or the Economine process, Sulfinol process, Propylene Carbonate process, and Selexol process. In the second part of the paper we will present solid gas treating processes including Iron Sponge, Mol Sieves, and in the third part of the paper we will present a description of Proprietary processes with direct reduction to sulfur including the Townsend process, Stretford process, and the Taxahax process.
The Ethanolamine sweetening process has been widely accepted in removing CO₂ and H₂S from natural gas streams. Monoethanol Amine has been used for many years and lately Diethanol Amine has come into favor with the gas treating industry. Since both processes use essentially identical equipment we will describe the process and then indicate differences in design calculations between the two.

The reader is referred to papers by Gene Goar, Charles Perry, Ward Rosen, and Allen Shell for more information on the Amine process. Jefferson Chemical has published brief papers on design and operating considerations for amine treaters.

A. DESCRIPTION OF PROCESS:

Referring to Fig.1 - 1 Natural Gas containing CO₂ and H₂S is contacted in countercurrent gas-liquid absorption process in a trayed or packed tower to provide intimate contact for a chemical reaction between H₂S - CO₂ (acid gas) and Amine. Good design practice dictates a scrubber on the inlet gas to remove entrained liquids including distillate and water from the gas before it enters the contactor. Also a scrubber on the outlet gas to recover any Amine solution carried over from the contactor should be provided.

Rich Amine from the bottom of the contactor is heated in a rich-lean Amine exchanger and then flashed at reduced pressure to remove part of the acid gas. The partially stripped solution is then passed to the stripper tower in which the solution is denuded of acid gas by steam stripping. Acid gases are concentrated in the overhead accumulator and disposed of either by burning in a flare, reboiler, or other incineration device. In cases of high sulfur volume, the acid gases may be processed for sulfur recovery. Lean amine solution from the bottom of the reboiler is exchanged with rich amine in the solution exchangers, then pumped in multi-stage pumps back to the contactor to complete the process loop.

Inasmuch as a clean solution is a key to the success of a treating system, good filtration is essential. Activated carbon filters have been found to provide the best and most economical filtration. Also MEA solutions may be reclaimed in a side stream reclaimer. DEA solutions cannot be reclaimed due to the high boiling point of DEA. (Our experience has not indicated reclaiming is necessary for DEA solutions.)
B. DESIGN CALCULATIONS:

1. Contactor

Pressure and gas volume are major influences on the contactor sizing. Referring to Figure 1-2 it is obvious that to minimize the size of the contactor, pressure should be at maximum practical.

2. Circulation

Circulation of amine solution is determined by the acid gas content of the inlet gas, strength of the solution, and the type of amine to be used. In Figures 1-3, we have provided quick determination of circulations required as a function of gas volume, acid gas content, and type of amine.

Amine circulation can be calculated by the following formulas:

\[
\text{For MEA} \quad \text{GPM} = \frac{Q \times X}{Z} \\
\text{For DEA} \quad \text{GPM} = \frac{23.0 \times Q \times X}{Z} \\
\]

where \( Q \) = Gas to be processed, MMSCFPD
\( X \) = Acid gas content, volume percent
\( Z = \frac{\text{Mol} \% \text{CO}_2 + \text{Grains H}_2\text{S}}{625} \)

The above formulas are based on mol loadings of 0.33 Mol/Mol for MEA and 0.7 Mol/Mol for DEA

3. Heat Loads

Estimates of heat loads and surface areas to be required are shown below as a function of circulation rates for amine solutions:

<table>
<thead>
<tr>
<th></th>
<th>Duty-BTU/HR</th>
<th>Area-Sq.Ft.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reboiler(Direct Fired)</td>
<td>72,000 x GPM</td>
<td>11.30 x GPM</td>
</tr>
<tr>
<td>Solution Exchangers</td>
<td>45,000 x GPM</td>
<td>11.25 x GPM</td>
</tr>
<tr>
<td>Solution Coolers</td>
<td></td>
<td></td>
</tr>
<tr>
<td>(Air Cooled)</td>
<td>15,000 x GPM</td>
<td>10.20 x GPM</td>
</tr>
<tr>
<td>Reflux Condenser</td>
<td></td>
<td></td>
</tr>
<tr>
<td>(Air Cooled)</td>
<td>30,000 x GPM</td>
<td>5.20 x GPM</td>
</tr>
</tbody>
</table>
4. **Power**

Power requirement for the process is largely for pumping of the amine solution. The table below gives power requirements as a function of amine solution circulation:

- **Solution pumps**  \( \text{GPM} \times \text{PSIG} \times 0.00065 = \text{HP} \)
- **Booster pumps**  \( \text{GPM} \times 0.06 = \text{HP} \)
- **Reflux pumps**  \( \text{GPM} \times 0.06 = \text{HP} \)
- **Aerial Cooler**  \( \text{GPM} \times 0.36 = \text{HP} \)

Normally, plant lighting and control power will add 2 - 5 KW depending on plant size.

**C. COST:**

1. **Capital Investment**

Capital investment for plants to use Monoethanol Amine processes can be broken into two segments. First, the contactor investment is a function of pressure and size and is presented on Fig. 1 - 4. Also on Fig. 1-4 is a cost for the regenerator unit as a function of amine circulation. In these curves it is assumed that the regenerator equipment will be skid-mounted and assembled in a shop rather than a field construction. The cost figures do not include transportation or foundations and piping required at the location.

2. **Operating Expenses**

Operating expenses for the amine unit will be largely operating labor and power. Typical operating expenses for a treating unit are indicated below: (Basis: small, unattended plant with 25 gpm DEA circulation, 3.0 MMCFD Gas)

- Operating Labor: 1.33
- Power: 0.33
- Chemical & Supplies: 0.11
- Supervision: 0.62
- Depreciation: 0.70
- Insurance and Taxes: 0.45
- Overhead (G&A): 0.57
- Repair and Maintenance: 0.22

**Total Cost**: \( - - - 4.38\$\text{/MCF} \)

**C. GRAPHS AND TABLES**
<table>
<thead>
<tr>
<th>Maximum Circulation (GPM)</th>
<th>Still Diameter (Inches)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>18</td>
</tr>
<tr>
<td>25</td>
<td>24</td>
</tr>
<tr>
<td>50</td>
<td>30</td>
</tr>
<tr>
<td>75</td>
<td>36</td>
</tr>
<tr>
<td>100</td>
<td>42</td>
</tr>
<tr>
<td>150</td>
<td>48</td>
</tr>
<tr>
<td>200</td>
<td>60</td>
</tr>
<tr>
<td>300</td>
<td>72</td>
</tr>
</tbody>
</table>

10.6 gpm / ft².
<table>
<thead>
<tr>
<th>Maximum Circulation (GPM)</th>
<th>Surge Tank Diameter (Inches)</th>
<th>Surge Tank Length (Inches)</th>
</tr>
</thead>
<tbody>
<tr>
<td>15</td>
<td>24</td>
<td>72</td>
</tr>
<tr>
<td>25</td>
<td>42</td>
<td>96</td>
</tr>
<tr>
<td>50</td>
<td>48</td>
<td>144</td>
</tr>
<tr>
<td>100</td>
<td>60</td>
<td>180</td>
</tr>
<tr>
<td>150</td>
<td>72</td>
<td>240</td>
</tr>
<tr>
<td>200</td>
<td>84</td>
<td>240</td>
</tr>
<tr>
<td>300</td>
<td>84</td>
<td>360</td>
</tr>
<tr>
<td>Maximum Circulation (GPM)</td>
<td>Reflux Accumulator Diameter (Inches)</td>
<td>Length (Inches)</td>
</tr>
<tr>
<td>---------------------------</td>
<td>-------------------------------------</td>
<td>-----------------</td>
</tr>
<tr>
<td>15</td>
<td>18</td>
<td>36</td>
</tr>
<tr>
<td>25</td>
<td>24</td>
<td>48</td>
</tr>
<tr>
<td>50</td>
<td>30</td>
<td>96</td>
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<tr>
<td>100</td>
<td>36</td>
<td>120</td>
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<tr>
<td>150</td>
<td>48</td>
<td>144</td>
</tr>
<tr>
<td>200</td>
<td>60</td>
<td>168</td>
</tr>
<tr>
<td>300</td>
<td>72</td>
<td>180</td>
</tr>
</tbody>
</table>
FIGURE 1-2
CONTACTOR CAPACITY

NATURAL GAS CAPACITY, MMSCF/D

OPERATING PRESSURE
Figure I-4
Amine Unit Costs

Delivered Cost for Each Component

Basis: Nelson Construction Index = 122
(1946 = 100)

Contactor Dia., Inches

Solution Rate GPM
SECTION 2

PROPRIETARY LIQUID PROCESSES

A. DIGLYCOLAMINE (ECONAMINE)

The Diglycolamine (DGA) process is carried out in equipment very similar to the equipment for Ethanolamine. The major difference being in the concentration of the Diglycolamine which is normally in the fifty to sixty percent concentration range. The DGA process does require a specialized purification unit on the side stream of the circulation.

Major advantages for the DGA process over the MEA process is a higher allowable concentration of DGA that is up to 60% with resulting higher acid gas loadings which leads to substantial savings in equipment size and cost, and utility cost. Optimum applicability is in the range of acid gas concentration from 1.5% to 8.0%.

B. SULFINOL (SHELL)

The Sulfinol process is unique in that it combines characteristics of a solvent process and an amine process. The high solveny of sulfinol for acid gases at high acid gas concentrations results in a greatly reduced circulation rate which is the basis of the economic advantages. The advantage disappears at low acid gas concentration. The sulfinol process has limitations which must also be considered. It has a great affinity for aromatics therefore pre-treatment facilities must be installed ahead of the acid gas if it is being fed to a sulfur recovery plant. Initial cost for the sulfinol process are quite high due to high sulfinol solution cost and royalties paid to Shell Development Company. The process has numerous advantages, especially its ability to selectively remove H₂S leaving up to 50% of CO₂ in the gas.

C. PROPYLENE CARBONATE

The Propylene Carbonate process has advantages over the Amine processes in that it has high acid gas loading and corresponding decrease in capital cost. Regeneration of the solvent is by pressure reduction requiring no steam regeneration. Disadvantages for the Propylene Carbonate process are similar to those of the Sulfinol process in that the solvent will retain heavy hydrocarbons so that acid gas must be treated to remove these hydrocarbons before being processed for sulfur recovery.
D. SELEXOL

Selexol process is a solvent process developed by Allied. Major advantages of the Selexol process are high solubility of H₂S relative to CO₂ making it possible to remove all H₂S leaving a certain amount of CO₂ in the residue gas. Selexol removes carbonyl sulfide, carbon disulfide, and mercaptan without degradation. The solvent will dehydrate natural gas if water is removed in the regeneration step. Disadvantages of the Selexol process is typical of other solvent processes, i.e., absorption of heavy hydrocarbons, expensive solvents, and royalty charges.

E. TYPICAL FLOW SHEETS

Typical flow sheets for Proprietary processes are presented as attachments.
REFERENCES


10. Maddox, Dr. R. N., and Burns, M. D., "Hot Carbonate - Another Possibility", The Oil and Gas Journal, October 9, 1967, pp. 167-173.


14. Maddox, Dr. R. N., and Burns, M. D., "Physical Solution is the Key to These Treating Processes", The Oil and Gas Journal, January 8, 1968, pp. 78-81.


SECTION 3
IRON SPONGE PROCESS

The iron sponge process is one of the oldest and simplest processes for treating sour gas. However, economics normally limits its application to gases containing less than 20 grs. H2S per 100 Scf. It may be used on low pressure gas, but is more successful on high pressure gas. The sponge must be moist to be reactive, and if the feed gas is dehydrated, it must be resaturated with water before entering the sponge bed.

The iron sponge material is oak wood shavings impregnated with a hydrated iron oxide. It is available as both the 9 lbs. and 15 lb. grade (based on iron content of 9 lbs./bushel and 15 lbs./bushel). A bushel of sponge will occupy one cubic foot of space when packed in a treating unit.

A. DESCRIPTION OF THE PROCESS:

An iron sponge treating unit is an open vessel sized for the proper bed volume and cross sectional area for the volume of gas to be processed. A screen, or coarse packing is placed in the bottom of the vessel to support the sponge. The remainder of the vessel is filled with iron sponge, leaving only a small void space at the top. Gas enters a top connection, passes downward through the bed, and then out the bottom connection. Hydrogen sulfide reacts with the iron oxide in the sponge to form iron sulfide, thus sweetening the gas. Carbon dioxide does not react with the sponge, and passes through the bed unchanged.

The iron sponge process is a batch process, thus requiring multiple towers, on interruption of flow when the bed is changed or regenerated.

B. DESIGN CALCULATIONS:

An iron sponge treating vessel must be designed based both on linear velocities through the bed and contact time. Once these parameters are determined, it is necessary to determine if the bed life will be satisfactory. The following is a description of these steps.

LINEAR VELOCITY:

The linear velocity through the bed should not exceed 10 ft. per minute based on the empty vessel and the gas at flow conditions. This will result in a pressure drop through the bed equivalent to approximately 1 - 2 Psi per foot of bed depth.
SPACE VELOCITY (or Contact Time):

The sponge bed should be of sufficient size to allow a minimum of one minute contact time based on the gas at flow conditions and the empty vessel. This will result in a space velocity of 60 actual cubic feet per hour per cubic foot of bed.

Figure 3-3 may be used to estimate the capacity of sponge vessels at various pressures.

In a typical sponge treating unit, the gas channels to some extent along the shell of the vessel. The H2S front moves through the bed as an inverted cone, and the greater the linear and space velocities, the steeper will be this cone. Thus if too high linear and space velocities are used, when the H2S breaks through around the vessel shell, there will be a large mass of unreacted sponge in the center.

BED LIFE:

Iron sponge theoretically can pickup 0.56 lbs. of sulfur per pound of iron oxide. For design purposes, this must be derated by about 25%. The following formula can be used to determine bed life:

\[
\text{MMcf/Change} = \frac{4.69}{\text{Bu. per Charge}} \times \frac{\text{Gr./100 Scf of H2S}}{
\]

The above formula is for 15 lbs. sponge. For 9 lb. sponge, the constant \( 2.80 \) should be substituted for 4.69.

AIR REGENERATION:

Iron sponge may be regenerated with air, either batch wise or continuously with some limitations. In batch regeneration, a spent sponge bed is isolated and depressurized. A blower is used to circulate gas through the bed at atmospheric pressure. Air is injected into the gas slowly over a 24 hour period, allowing the oxygen content in the gas to slowly increase to 8%. The bed life after each regeneration will be approximately 60% of the previous bed life, resulting in only about 2 regenerations being feasible. After a bed has been regenerated two or three times, elemental sulfur formed on regeneration builds around the sponge making it extremely difficult to remove the spent sponge.
C. **COSTS**

Capital costs of approximately $6000 per MMcfd capacity (at 1 - 2 MMcfd) to $2000 per MMcfd capacity (at 30 MMcfd) may be used to estimate installed costs for 1000 Psi units.

For operating costs, $0.01/Mcf per for each 3 to 5 gr./100 Scf in the inlet gas will cover the costs of sponge, labor, and amortization of cost of equipment.

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**REFERENCES**


2. Taylor, D. K., "How to Desulfurize Natural Gas," A series in *The Oil and Gas Journal*, November 5 and 19, and December 3 and 10, 1956

3. Maddox, Dr. R. N., and Burns, M. D., "Iron Oxide Process Design Calculations", *The Oil and Gas Journal*, August 12, 1968


# TABLE 1

**BED CAPACITIES FOR COMMON SIZE SPONGE TREATERS**

<table>
<thead>
<tr>
<th>Vessel Dia., Inches</th>
<th>Bed Capacity, Bushels (or Cu. Ft.) For Bed Depth Indicated</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>10'</td>
</tr>
<tr>
<td>24&quot; OD</td>
<td>30</td>
</tr>
<tr>
<td>30&quot; ID</td>
<td>50</td>
</tr>
<tr>
<td>36&quot; ID</td>
<td>70</td>
</tr>
<tr>
<td>42&quot; ID</td>
<td>96</td>
</tr>
<tr>
<td>48&quot; ID</td>
<td>126</td>
</tr>
<tr>
<td>54&quot; ID</td>
<td>160</td>
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<td>60&quot; ID</td>
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<tr>
<td>66&quot; ID</td>
<td>238</td>
</tr>
<tr>
<td>72&quot; ID</td>
<td>284</td>
</tr>
<tr>
<td>84&quot; ID</td>
<td>384</td>
</tr>
<tr>
<td>96&quot; ID</td>
<td>500</td>
</tr>
</tbody>
</table>
FIGURE 3-2 CAPACITY OF IRON OXIDE FOR SOUR GAS BETWEEN RECHARGES

BASED ON: 15 LB. Fe₂O₃/BUSHEL
0.56 LB. S PICKUP/LB. Fe₂O₃
75% BED SATURATION

IRON SPONGE - BUSHELS PER CHARGE

FORMULA FOR GRAPH:

\[
\text{MMCF/CHARGE} = 4.69 \frac{\text{BU.}}{\text{GR./100 SCF}}
\]

CRP
2-28-70
FIGURE 3-3 LINEAR AND SPACE VELOCITIES FOR IRON SPONGE TREATERS

BASED ON:
- MAXIMUM LINEAR VELOCITY: 10 FT./MIN.
- MAXIMUM SPACE VELOCITY: 60 CU. FT. PER HOUR PER CU. FT. OF BED FOR 10 GR./100 H₂S CONTENT

TREATMENT CAPACITY (MMCF) vs. TREATMENT TIME (HOURS) vs. TREATMENT PRESSURE (PSIG)
SECTION 4

PROPRIETARY SOLID PROCESS
(Molecular Sieve)

A. DESCRIPTION OF THE PROCESS:

In the molecular sieve process, the sour gas passes through one of two, three, or four fixed beds of molecular sieves where the hydrogen sulfide, along with water and organic sulfur compounds, are removed from the gas by a process similar to adsorption. Whenever the bed becomes saturated with hydrogen sulfide, the main gas flow is switched to another bed which is freshly regenerated. This cycling of beds continues on a regular cycle, based either on time, or on break-through of hydrogen sulfide.

A side stream of the sweet gas (usually approximately 20%) is heated to approximately 600° to 700° F. and passed through the last fouled bed to regenerate it. The hot regeneration gas is then cooled and processed by a wet process (usually amine or sulfinol) to remove the hydrogen sulfide from the regeneration gas. After the regeneration gas is sweetened, it rejoins the main gas stream downstream of the sieve beds.

B. ADVANTAGES AND APPLICATIONS:

The molecular sieve process may be used to selectively remove hydrogen sulfide in the presence of carbon dioxide. Thus, if it is not necessary to remove carbon dioxide from a gas stream, the molecular sieve process may offer a lower capital cost, lower fuel requirements, and/or lower operating expenses. Molecular sieves will also remove organic sulfur compounds.

 offsetting these advantages, the necessity of a separate regeneration gas treating unit makes this process somewhat complicated and less applicable to small streams. While the sieves can selectively remove hydrogen sulfide, the presence of carbon dioxide does adversely affect the capacity of the sieves for H2S. Also, sieves are subject to attack by contaminants occasionally found in gas, and care must be taken to see that the incoming gas is free of contaminants.
REFERENCES


SECTION 5

PROPRIETARY PROCESSES WITH DIRECT REDUCTION TO SULFUR

A. TOWNSEND PROCESS:

In the Townsend Process, an SO₂ rich organic solvent, such as triethylene glycol, contacts the sour gas, and the hydrogen sulfide reacts with the SO₂ to form elemental sulfur and water. The glycol solvent also absorbs additional water vapor from the gas. The solvent, containing element sulfur and water, is heated to melt the sulfur and distill out the water. The liquid sulfur is decanted, and a portion is burned to form SO₂, which is absorbed in the lean solvent, and recycled to the contactor.

At present, various types of catalysts may be used to catalyze the H₂S - SO₂ reaction, and other solvents are being investigated.

REFERENCE

B. **STRETFORD PROCESS:**

The Stretford Process is an absorption process, in which the H₂S rich solution is air blown to reduce the H₂S to finely divided elemental sulfur particles, which are removed by centrifuge or filtration. The air blowing and reduction of the H₂S regenerates the solvent for recirculation.

The solvent used in the Stretford Process is an aqueous solution of sodium carbonate and anthraquinone disulfonic acid with a sodium metavanadate activator.

The chemicals in the solvent are all quite stable, but the presence of air and sulfur does cause the formation of a small amount of thiosulfate in the form of a sodium salt. The process produces good purity sulfur, and can be used for low H₂S: CO₂ ratio streams.

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**REFERENCES**


C. TAKAHAX PROCESS:

The Takahax Process is similar to the Stretford Process, in that a special solvent removes the H₂S from the gas, and the rich solution is air blown to reduce the H₂S to elemental sulfur and regenerate the solvent.

In the Takahax Process, the solvent is an alkaline solution of 1,4-naphtaquinone, 2-sulfonate sodium, and a catalyst. Sulfur formed in the solution during regeneration is removed by filtration.

REFERENCES