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FUNDAMENTALS OF GAS TREATING

MANUAL FOR DESIGN OF GAS TREATERS

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# F U N D A M E N T A L S   O F   G A S   T R E A T I N G

## INTRODUCTION

In this paper it is our intention to review the fundamentals of gas treating by liquid and solid processes. In Section 1 we describe the Ethanolamine Process for treating gas to remove acid gas constituents. In Section 2 Iron Sponge as a solid gas treating process is described. In Section 3 we present a description of Proprietary liquid processes, and in Section 4 Proprietary solid processes are discussed.

Section 5 presents proprietary processes that incorporate sulfur recovery. A brief description of the Claus Recovery Process and the Recycle Selectox Sulfur Recovery Process are presented in Section 6.

References and selected reading materials are listed in Section 7.

## FACTORS TO CONSIDER IN SELECTING A TREATING PROCESS:

1. Treating Pressure (Inlet Gas Pressure)
2. Acid Gas Content ( $\text{CO}_2 + \text{H}_2\text{S}$ )
3. COS/CS<sub>2</sub> Content
4. Treated Gas Specifications
5. Availability of Power
6. Environmental Constraints

## SECTION 1

### MONOETHANOL/DIETHANOL AMINE PROCESS

The Amine sweetening process has been widely accepted in removing CO<sub>2</sub> and H<sub>2</sub>S 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 Gas Purification, by Kohl and Riesenfield, plus papers by Gene Goar, Charles Perry, Ward Rosen, and Allen Shell for more information on the Amine Process.\*\* Jefferson Chemical has published brief design and operating considerations for amine treaters.

#### A. DESCRIPTION OF PROCESS

Referring to Figure 1-1, Natural Gas containing CO<sub>2</sub> and H<sub>2</sub>S is contacted in countercurrent gas-liquid absorption process in a trayed or packed tower to provide intimate contact for a chemical reaction between H<sub>2</sub>S - CO<sub>2</sub> (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.

\*The Monoethanol Amine process was first commercially applied as the "Girbitol" process in 1930.

\*\*A complete set of references is listed in Section 7.

Rich Amine from the bottom of the contactor is flashed at reduced pressure to remove entrained gases, including part of the acid gas and then heated in a rich-lean Amine exchanger. The solution is then fed to the stripper tower where the solution is regenerated and denuded of acid gas by steam stripping. Acid gases are concentrated in the overhead accumulator and disposed of by burning in a flare, reboiler, or other incineration device. In cases where sulfur exceeds 2-5 tons per day, 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 reclainer. DEA solutions cannot be reclaimed due to the high boiling point of DEA. (Our experience has indicated reclaiming is not necessary for DEA solutions.)

#### B. DESIGN CALCULATIONS

##### 1. Input Data

- a. Gas Flow - Minimum, Normal, Maximum
- b. Gas Pressure - Minimum, Normal, Maximum
- c. Gas Temperature - Minimum, Normal, Maximum
- d. Gas Composition - Sp Gr, H<sub>2</sub>S, CO<sub>2</sub>, COS, CS<sub>2</sub>, H<sub>2</sub>O
- e. Specs for Treated Gas
- f. Utilities Available

## 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 Figure 1-2, 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:

$$\text{For MEA: } \text{GPM} = 41.0 \times QX/Z$$

$$\text{For DEA: } \text{GPM} = 45.0 \times QX/Z \text{ (conventional)}$$

$$\text{GPM} = 32.0 \times QX/Z \text{ (high-load)}$$

$$\text{For DGA: } \text{GPM} = 77.0 \times QX/Z$$

Where  $Q$  = Gas to be processed, MMscfd

$X$  = Acid gas content, volume percent

$$= \text{Mol \% CO}_2 + \frac{\text{Grains H}_2\text{S}}{632}$$

$Z$  = Amine concentration, wt. %

The above formulas are based on mol loadings of 0.33 Mol/Mol for MEA, 0.5 Mol/Mol for conventional DEA, 0.7 Mol/Mol for high-load DEA, and 0.3 Mol/Mol for DGA.

## 3. Contactor

Pressure and gas volume are major influences on the contactor sizing. Referring to Figure 1-3, the size of the contactor may be determined as a function of pressure and gas flow.

4. Regeneration Vessels - Table 1-2

- a. Still
- b. Surge tank
- c. Reflux Accumulator
- d. Flash-Skimmer Tank
- e. Carbon Filter

5. Heat Loads

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

	<u>Duty-Btu/Hr.</u>	<u>Area-Sq. Ft.</u>
a. Reboiler (Direct Fired)	72,000 x GPM	11.30 x GPM
b. Solution Exchangers	45,000 x GPM	11.25 x GPM
c. Solution Coolers (Air Cooled)	15,000 x GPM	10.20 x GPM
d. Reflux Condenser (Air Cooled)	30,000 x GPM	5.20 x GPM

6. Power Required

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

$$HP = \frac{GPM \times \Delta P}{1713 \text{ (efficiency of pump)}}$$

Solution Pumps	GPM x Psig x 0.00065	= HP
Booster Pumps	GPM x 0.06	= HP
Reflux Pumps	GPM x 0.06	= HP
Aerial Cooler	GPM x 0.36	= HP

C. COST

1. Capital Investment

Capital investment for plants to use either MEA or DEA process can be broken into two segments. First, the contactor investment is a function of pressure, gas flow, and amine circulation. The cost for the regenerator unit is a function of amine circulation. Figure 1-4 presents investment data for battery limit gas treating plants as a function of amine circulation. It is assumed that the regenerator equipment will be skid-mounted and assembled in a shop to minimize field construction. The cost figures include transportation, foundations, and piping required at the location as a turn key installation.

2. Operating Expenses

Operating expenses for the amine unit will be largely operating labor, power and fuel gas. Typical operating expenses for a treating unit are presented Table 1-1.

TABLE 1-1

## AMINE UNIT OPERATING COSTS

Basis: Treating plant with 40GPM circulation; treating 3MMcfd gas  
with 10% acid gas: January, 1981, cost index

	<u>\$/Mcf</u>	<u>\$/Year</u>
Operating Labor (2 @ 22,000).....	4.07	44,000
Supervision (1 @ 35,000).....	3.24	35,000
Employee Benefits @ 35% Payroll.....	2.59	28,000
Utilities .....	2.78	30,000
Chemicals & Supplies.....	1.39	15,000
Repair Materials & Labor @ 4% Investment.....	4.44	48,000
Direct Overhead @ 5% Investment.....	5.57	60,000
Corporate Overhead @ 3.5% Investment.....	3.89	42,000
Depreciation ( 10 Year Straight Line).....	11.11	120,000
Interest ( @ 20%, first year).....	22.22	240,000
Insurance & Taxes @ 2% Investment.....	2.22	24,000
	<u>63.52</u>	<u>686,000</u>

Plant Investment Estimate \$1,200,000

**TABLE I - 2**  
**REGENERATION VESSEL SIZES**  
**(Inches)**

K-10

Solution Circl. Rate GPM	Still Diameter	Surge Tank		Reflux Accum.		Flash Tank		Carbon Filter	
		Diam.	Length	Diam.	Length	Diam.	Length	Diam.	Length
10	16	24	72	16	36	24	72	16	84
25	24	42	96	24	48	42	96	24	84
50	30	48	144	30	96	48	144	36	96
100	42	60	192	42	96	60	192	48	96
200	60	84	288	60	96	84	288	60	96
300	72	84	384	72	96	84	384	72	96
400	84	96	384	84	96	96	384	84	96

K-6

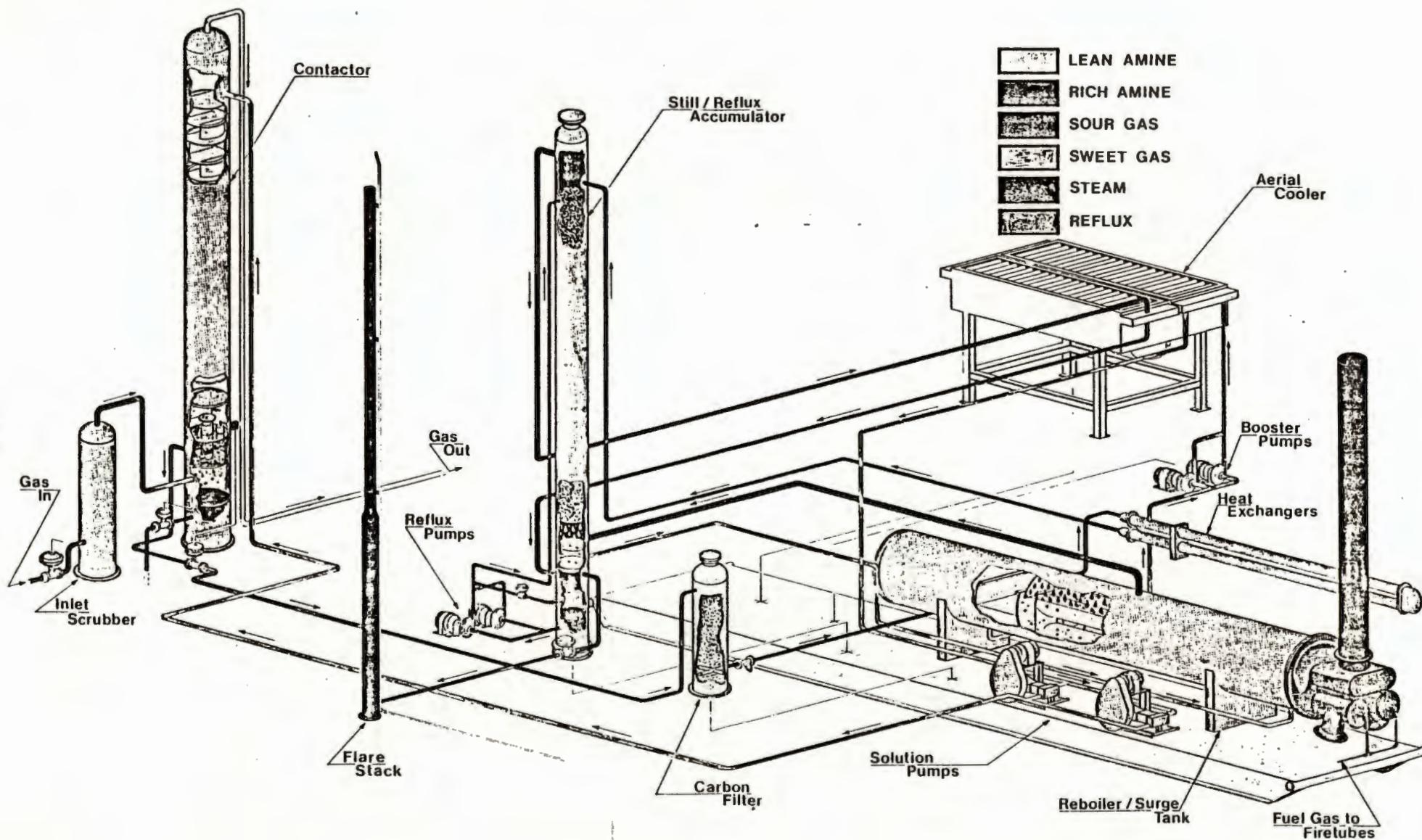
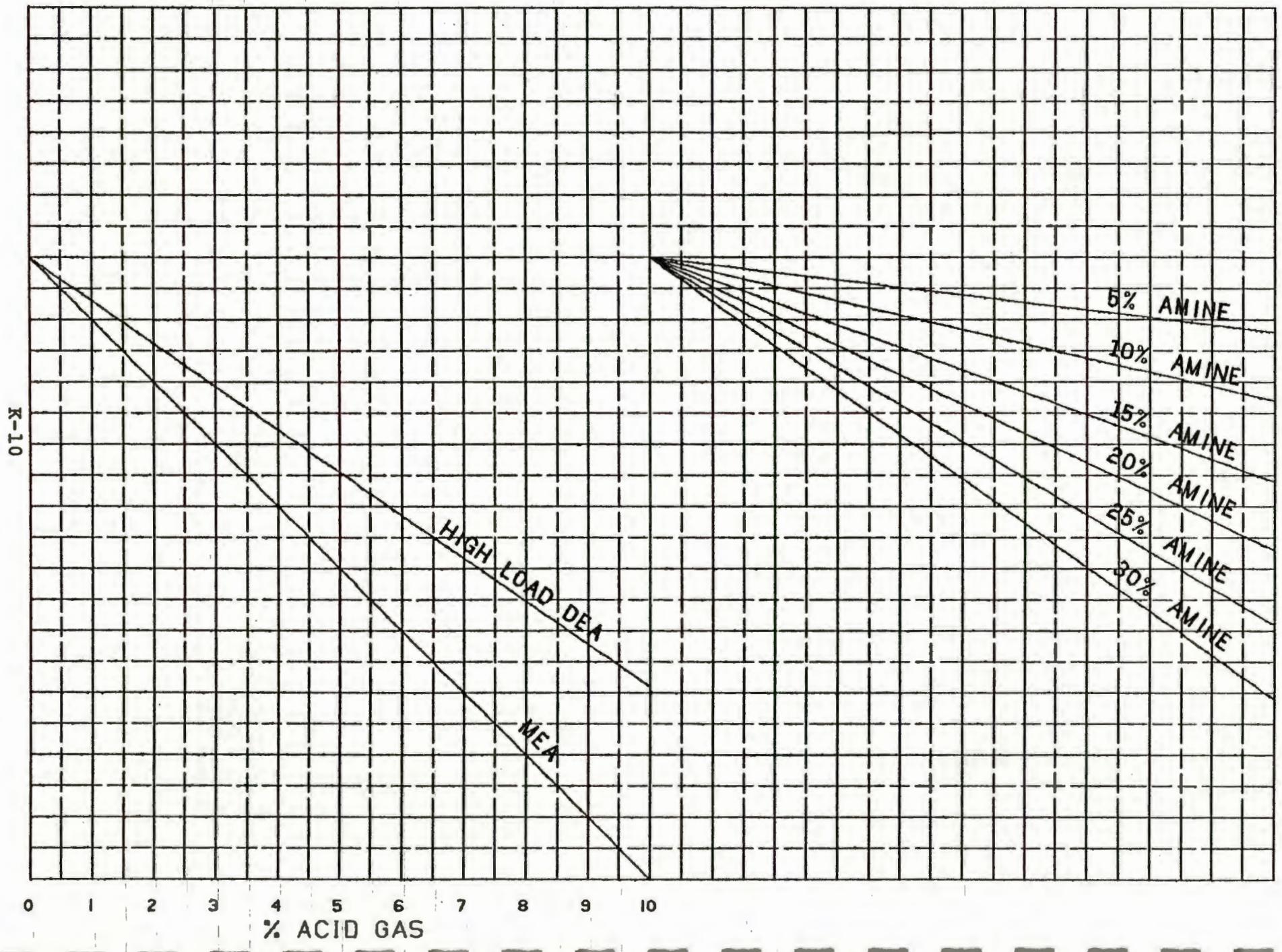
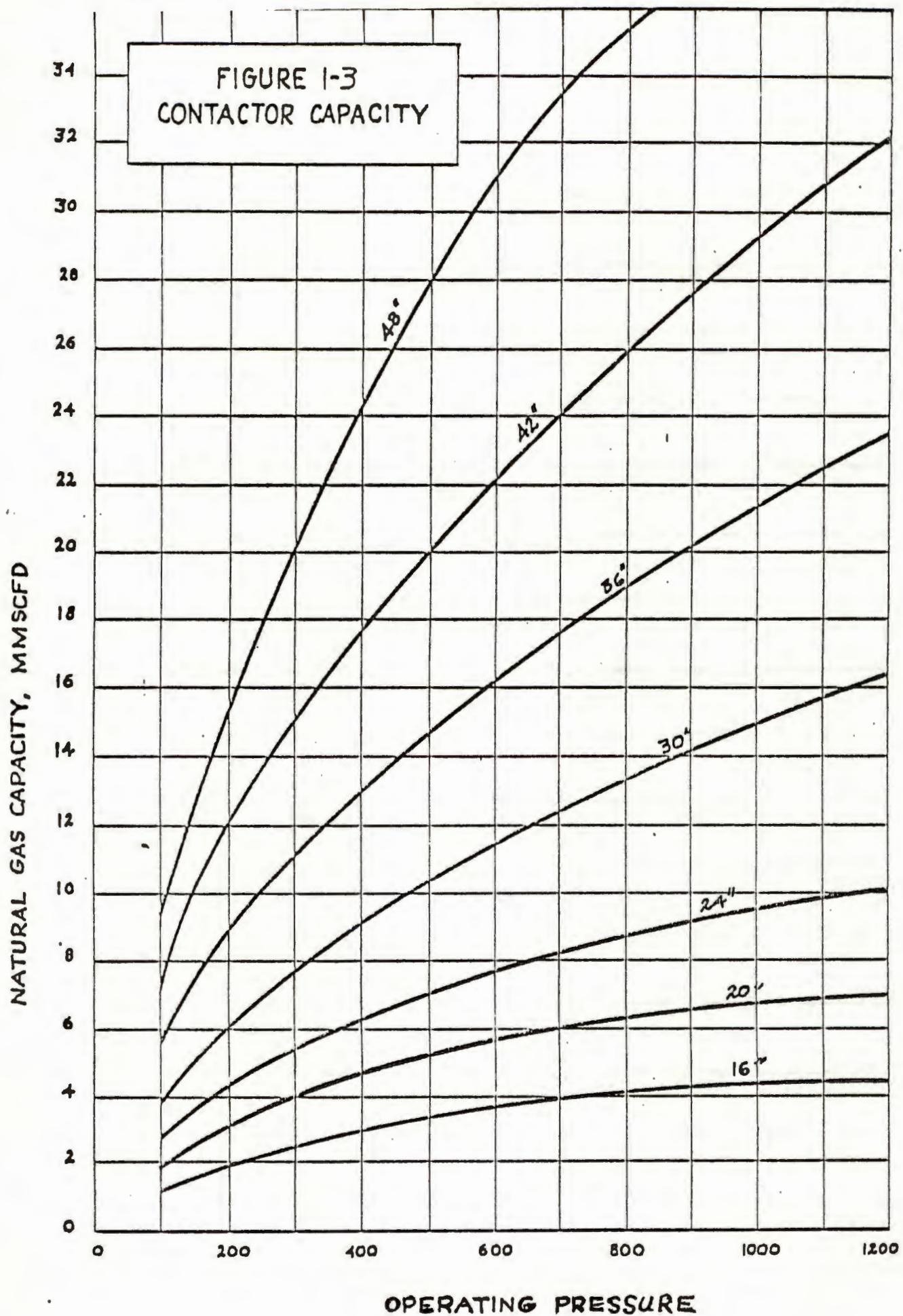


FIGURE I-2  
AMINE CIRCULATION REQUIRED

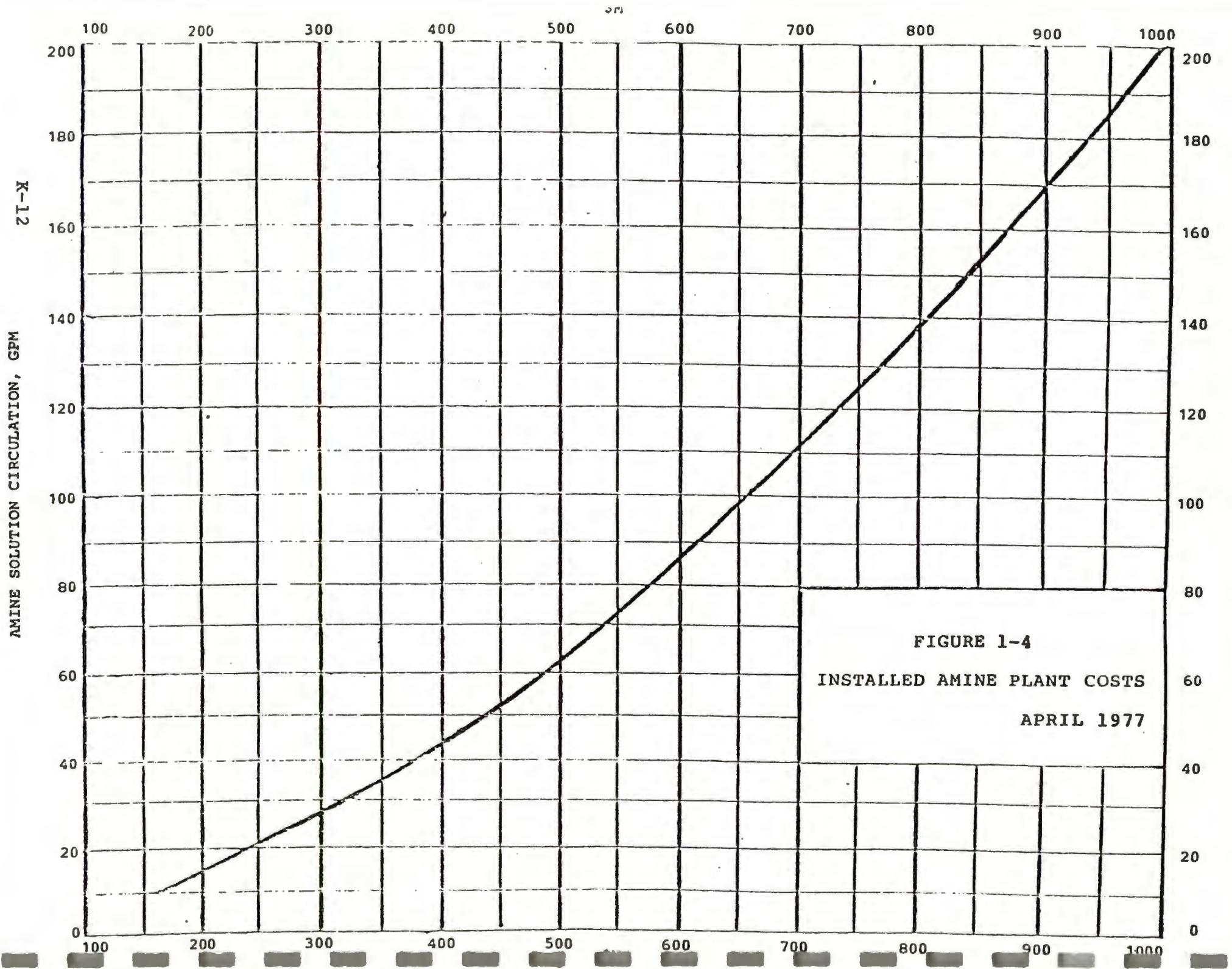
GPM CIRCULATION REQUIRED PER MMCFD

0 1 2 3 4 5 6 7 8 9 10





OPERATING PRESSURE



## SECTION 2

### 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. H<sub>2</sub>S 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 lb. 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.

1. Linear Velocity

The superficial linear velocity through the bed should not exceed 10 ft. per minute based on the empty vessel cross sectional area and the gas volume corrected to ACFM at flowing temperature and pressure. Pressure drop through the bed should be approximately 1 - 2 Psi per foot of bed depth.

2. Space Velocity (or Contact Time)

The sponge bed should be of sufficient size to allow a maximum space velocity of 6.0 ACFH per cubic foot of bed for each grain of H<sub>2</sub>S in the inlet gas.

Figure 2-3 may be used to estimate the capacity of sponge vessels at various pressures.

### Space Velocity (cont'd)

In a typical sponge treating unit, the gas channels to some extent along the shell of the vessel. The H<sub>2</sub>S 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 H<sub>2</sub>S breaks through around the vessel shell, there will be a large mass of unreacted sponge in the center.

### 3. 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:

$$\text{MMcf/Change} = \frac{\text{Bu. Per Charge}}{\text{Gr./100 of H}_2\text{S}}$$

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

#### 4. 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 depressured. 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. COST

##### 1. Capital Investment

Capital costs of approximately \$15,000.00 per Mmcfd capacity (up to 20 MMcfd) to \$10,000.00 per MMcfd capacity (above 30 MMcfd) may be used to estimate installed costs for 1000 Psi units.

##### 2. Operating Expenses

For operating costs, 0.12/Mcf for each 5 gr./100 Scf in the inlet gas will cover the costs of sponge, labor and amortization of cost of equipment.

Table 2-1

BED CAPACITIES FOR COMMON SIZE SPONGE TREATERS

Vessel Dia., Inches	Bed Capacity, Bushels (or Cu. Ft) For Bed Depth Indicated				
	<u>10'</u>	<u>12'</u>	<u>15'</u>	<u>18'</u>	<u>20'</u>
24" OD	30	34	44	52	58
30" ID	50	60	74	88	98
36" ID	70	84	104	118	140
42" ID	96	116	144	174	192
48" ID	126	150	190	226	252
54" ID	160	190	240	286	320
60" ID	196	236	294	354	392
66" ID	238	286	356	428	474
72" ID	284	340	424	410	568
84" ID	384	462	578	692	768
96" ID	500	600	752	902	1004

K-18

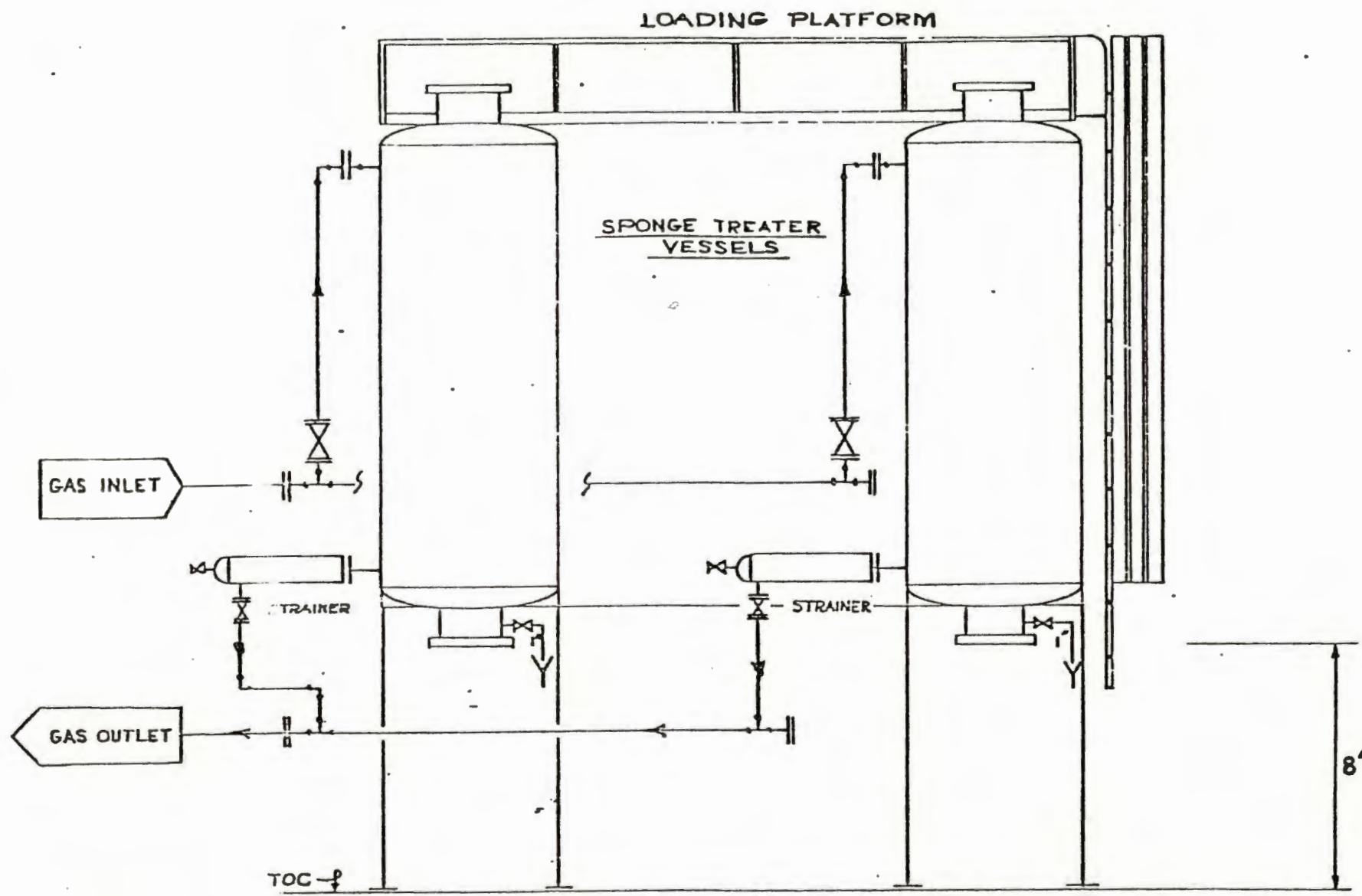
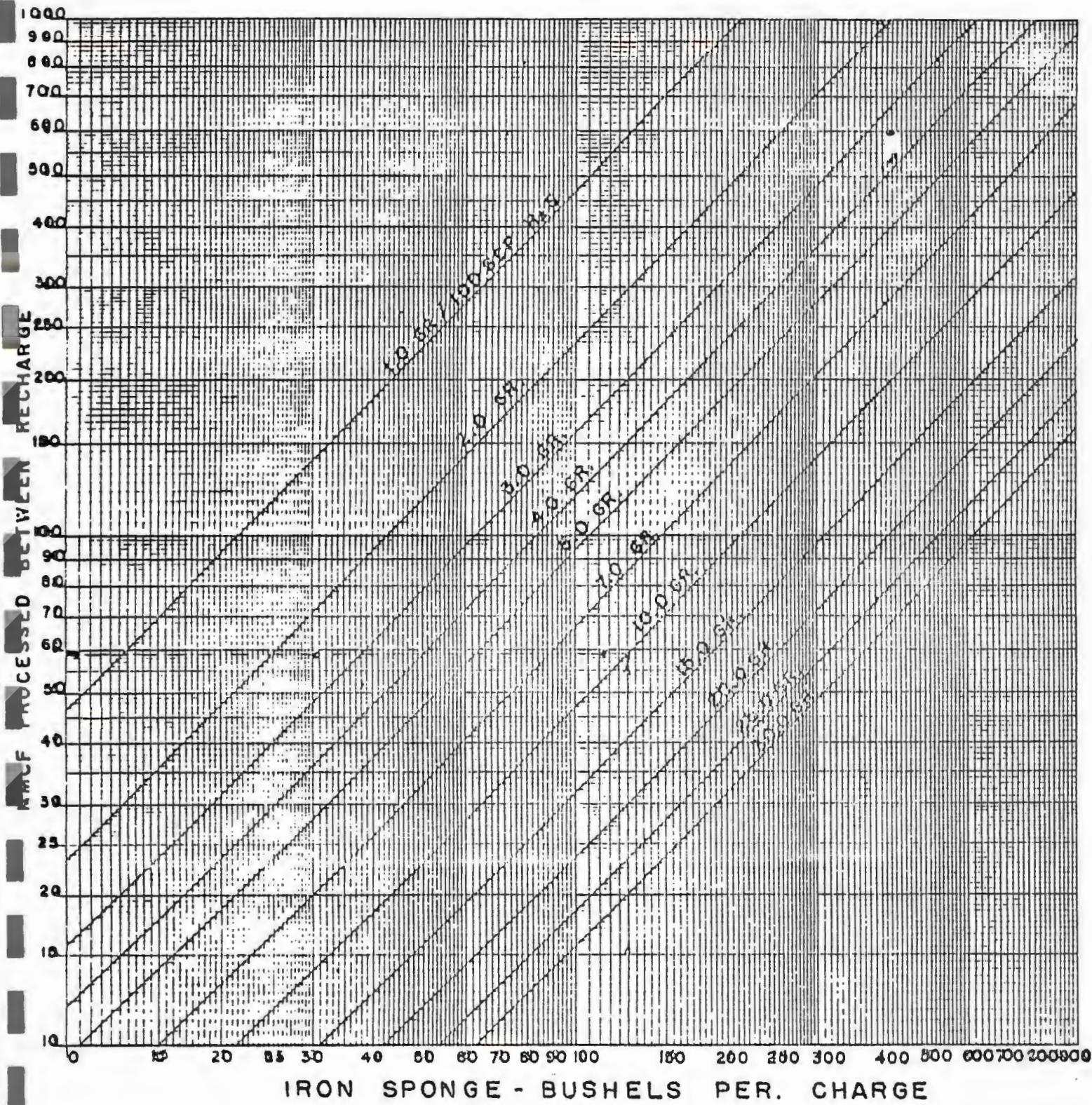


FIGURE 2-1  
FLOW DIAGRAM FOR IRON SPONGE TREATER PLANT

**FIGURE 2-2 CAPACITY OF IRON OXIDE FOR SOUH GAS  
BETWEEN RECHARGES**

BASED ON: 15 LB. FE<sub>2</sub>O<sub>3</sub>/BUSHEL  
0.56 LB. S PICKUP/LB. FE<sub>2</sub>O<sub>3</sub>  
75 % BED SATURATION



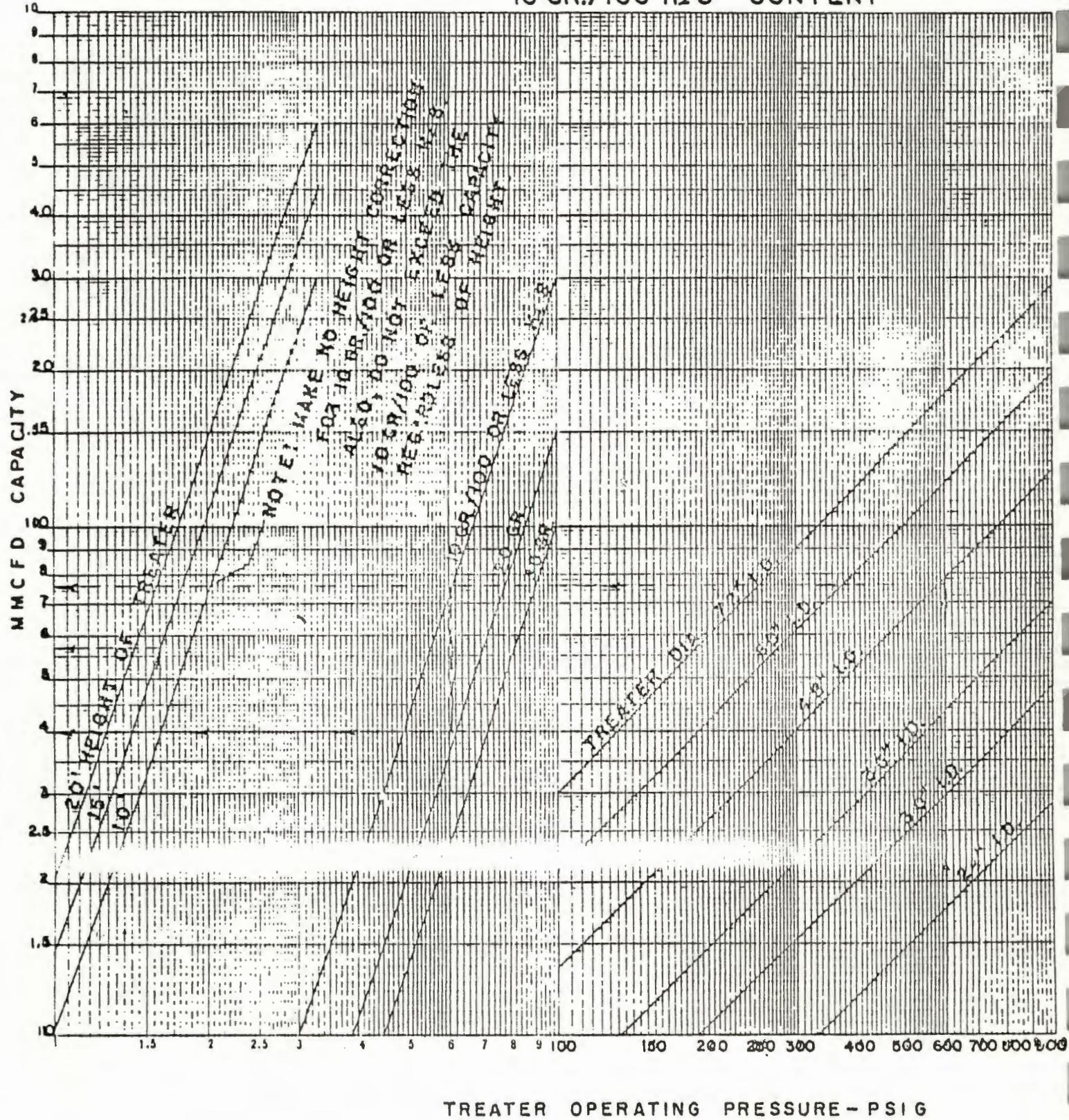
FORMULA FOR GRAPH:

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

K-19

## FIGURE 2-3 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 H<sub>2</sub>S CONTENT



## SECTION 3

### 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 solvency 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 is 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 H<sub>2</sub>S leaving up to 50% of CO<sub>2</sub> 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 H<sub>2</sub>S relative to CO<sub>2</sub> making it possible to remove all H<sub>2</sub>S, leaving a certain amount of CO<sub>2</sub> 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. MDEA

The ethanolamine amine process utilizing n-methyldiethanol amine "MDEA" is carried out in equipment very similiar to the equipment used for primary and secondary ethanolamines. The major difference being the concentration of the MDEA, the number of trays in the stripper and absorber, and the heat load for the process.

Major advantages of the MDEA process over MEA, or DEA, is the capacity of higher selectivity with respect to H<sub>2</sub>S in the presence of CO<sub>2</sub>. This can result in substantial savings in equipment size and cost, and utility cost depending on the desired and achievable selectivity.

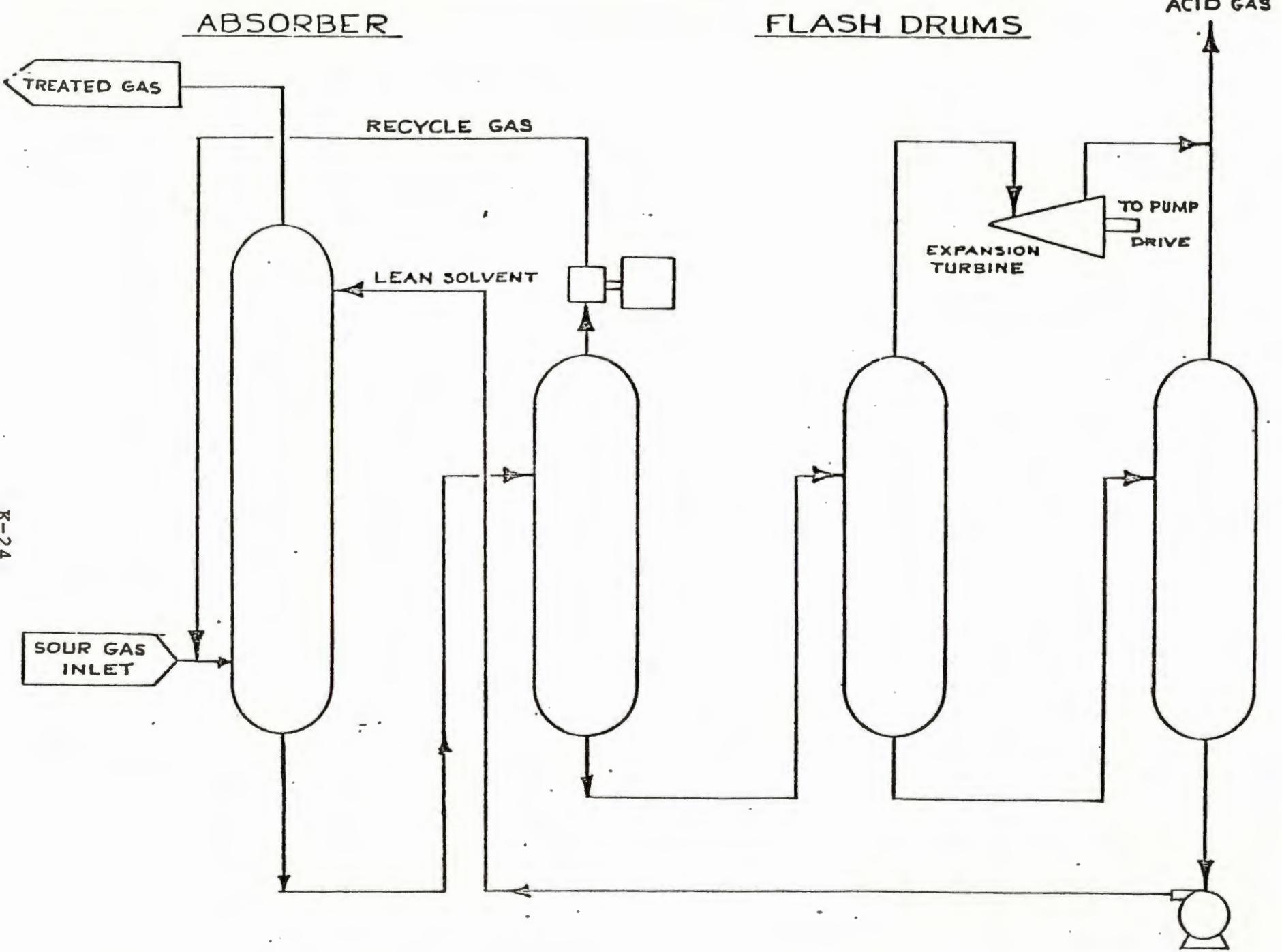
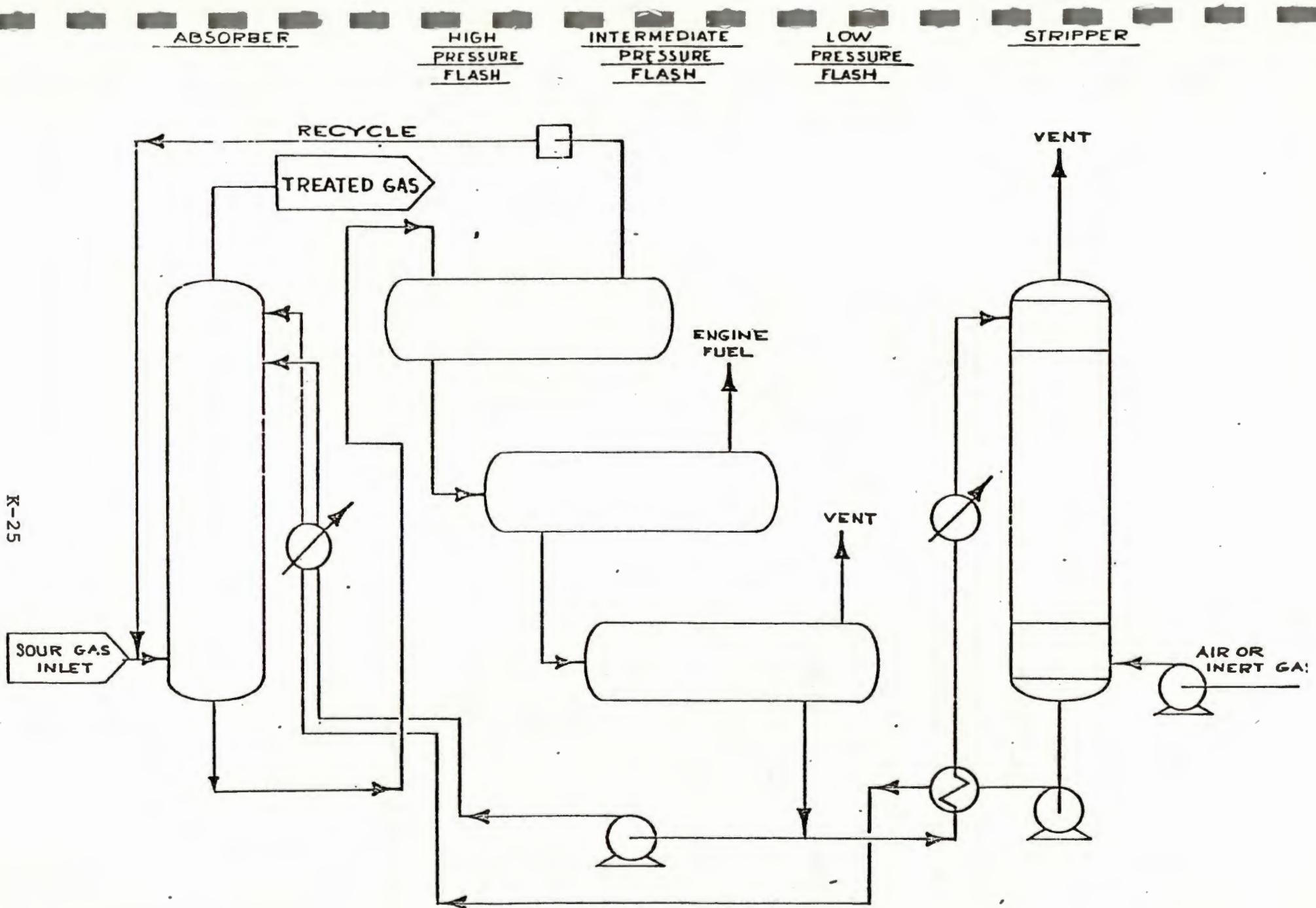


FIGURE 3-1  
PROPYLENE CARBONATE PROCESS



**FIGURE 3-2**  
**SELEXOL PROCESS**

## 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 breakthrough of hydrogen sulfide.

A side stream of the sweet gas (usually approximately 20%) is heated to approximately 600°F. 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 H<sub>2</sub>S. 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.

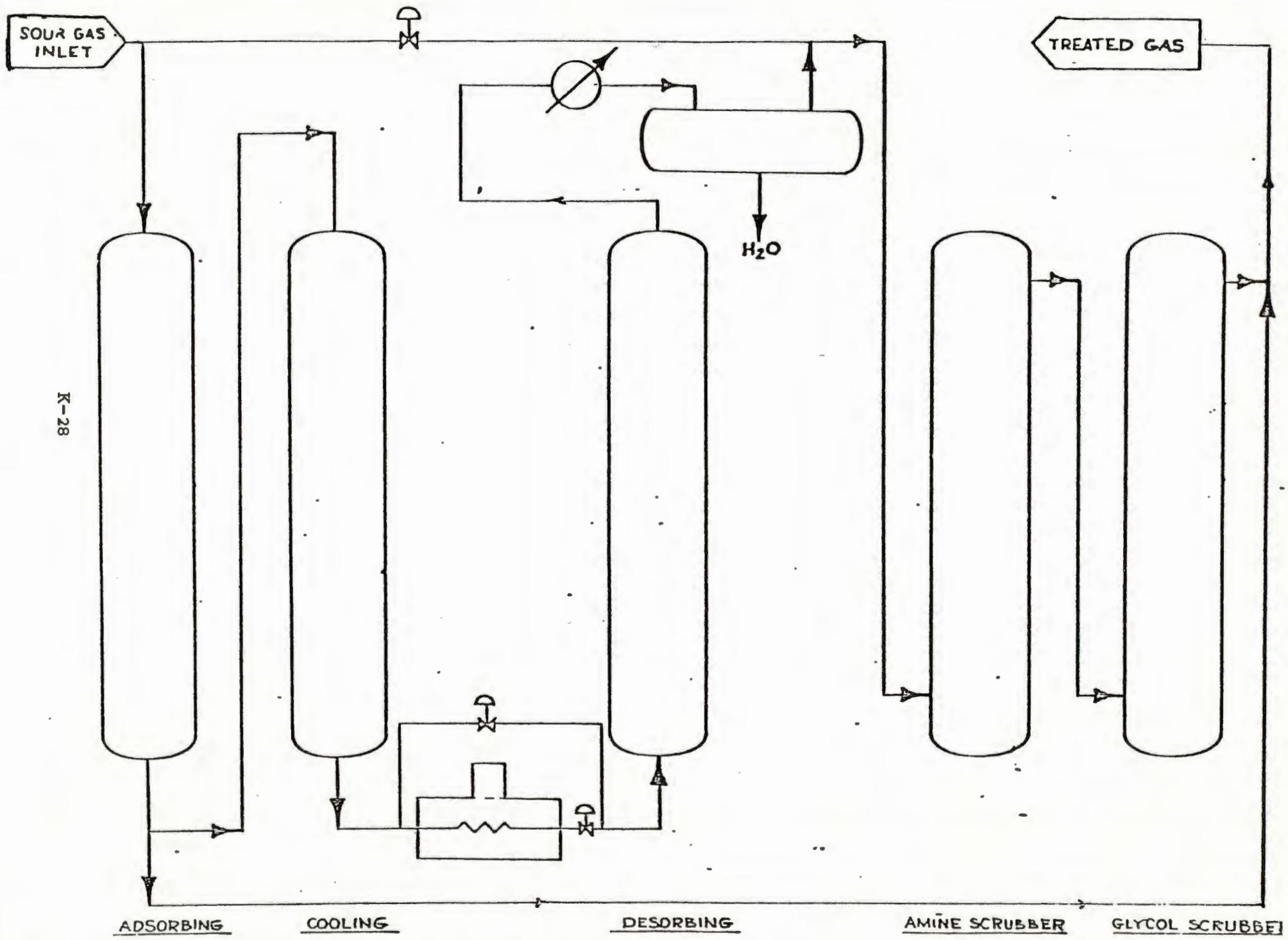


FIGURE 4-1  
NEW SEVERE ROSES

## SECTION 5

### PROPRIETARY PROCESSES WITH DIRECT REDUCTION TO SULFUR

#### A. TOWNSEND PROCESS

In the Townsend Process, an SO<sub>2</sub> rich organic solvent, such as triethylene glycol, contacts the sour gas, and the hydrogen sulfide reacts with the SO<sub>2</sub> to form elemental sulfur and water. The glycol solvent also absorbs additional water vapor from the gas. The solvent, containing elemental 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 SO<sub>2</sub>, which is absorbed in the lean solvent, and recycled to the contactor.

At present, various types of catalysts may be used to catalyze the H<sub>2</sub>S - SO<sub>2</sub> reaction, and other solvents are being investigated.

B. STRETFORD PROCESS

The Stretford Process is an absorption process, in which the H<sub>2</sub>S rich solution is air blown to reduce the H<sub>2</sub>S to finely divided elemental sulfur particles, which are removed by centrifuge or filtration. The air blowing and reduction of the H<sub>2</sub>S 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 H<sub>2</sub>S: CO<sub>2</sub> ratio streams.

C. TAKAHAX PROCESS

The Takahax Process is similar to the Stretford Process, in that a special solvent removes the H<sub>2</sub>S from the gas, and the rich solution is air blown to reduce the H<sub>2</sub>S to elemental sulfur and regenerate the solvent.

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

SECTION 6  
SULFUR RECOVERY PROCESS

A. CLAUS PROCESS FOR SULFUR RECOVERY

The most widely accepted process for recovering sulfur from acid gas streams is the modified Claus Process. The basic Claus reaction was disclosed in 1893:



For convenience in analyzing the steps in the plant operations, this reaction can be broken down into two steps:



The reader is referred to Gas Purification by Kohl and Riesenfeld plus papers by Gene Goar. A series of articles by Dr. Richard K. Kerr, Director of Applied Research and Development, Western Research and Development of Calgary, Alberta, has appeared in Energy Processing / Canada, which describes the Claus Process Thermodynamics and Kinetics.

1. DESCRIPTION OF PROCESS

Referring to Figure 6-1, acid gas from the amine treater containing H<sub>2</sub>S and CO<sub>2</sub> is pre-heated and injected into the reaction furnace along with sufficient air to burn 1/3 of the H<sub>2</sub>S. Sufficient resident time is provided in the reaction furnace to permit a part of the reaction between H<sub>2</sub>S and SO<sub>2</sub> to proceed. Indeed, approximately 65% of the recovered sulfur is formed in the reaction furnace. The reactants from the furnace are cooled in a waste-heat recovery boiler to approximately 1100°F.

The partially cooled reactants are split with a portion of the 1100°F. stream being cooled to 275°F. where molten sulfur is condensed. Unreacted H<sub>2</sub>S and SO<sub>2</sub> join a portion of the hot gas by-pass material for re-heat to approximately 600°F. for entry into the first catalytic reactor where a bauxite (alumina) catalyst is used to improve the reaction kinetics. The effluent from the No. 1 reactor is cooled in No. 2 condenser to drop out liquid sulfur, with unreacted H<sub>2</sub>S and SO<sub>2</sub> passing on in similar manner to reactors No. 2 and No. 3. Tailgas from the No. 4 condenser is normally sent to an incinerator where the last traces of H<sub>2</sub>S are burned to SO<sub>2</sub> and discharged in a tall stack.

Where very stringent environmental restrictions dictate a 99+% recovery of sulfur, the tailgas will be further processed in a "tailgas clean-up unit" such as Shell's SCOT Process or other other tailgas clean-up processes.

## 2. CAPITAL INVESTMENT

Capital investment for Claus Sulfur Plants is somewhat dependent on the percent of H<sub>2</sub>S in the acid gas from the sulfur plant. Figure 6-2 presents capital costs for a three-stage Claus unit plus incinerator presented in a recent publication by the EPA.\*

\* "An Investigation of the Best Systems of Emission Reduction for Sulfur Compounds from Crude Oil and Natural Gas Field Processing Plants"; W.O. Herring and R.E. Jenkins, Research Triangle Park, North Carolina.

Figure 6-3 presents investment costs for a two-stage Claus unit plus tailgas clean-up unit employing the Wellman-Lord process. Table 6-1 presents annual operating costs for Claus Sulfur Plants with incinerators. Table 6-2 presents operating costs for Claus Sulfur Recovery Unit with Wellman-Lord tailgas clean-up unit. These operating costs are also taken from the EPA publication.

#### B. RECYCLE SELECTOX SULFUR RECOVERY PROCESS

A new technology recovering sulfur from gas streams has been developed based on a new remarkable active Selectox catalyst which selectively oxidizes H<sub>2</sub>S to sulfur using air at low temperatures without forming SO<sub>2</sub> or oxidizing either hydrogen or light hydrocarbons. The process developed jointly by the Science and Technology Division of Union Oil Company of California and The Ralph M. Parsons Company has application in a wide range of H<sub>2</sub>S feed compositions, but especially shines in concentrations from 1 to 40% H<sub>2</sub>S.

##### 1. DESCRIPTION OF PROCESS

Referring to Figure 6-5 the acid gas is fed directly to the Selectox catalyst after being mixed with a stoichiometric amount of air. Feed temperature is about 350°F against 450°F for a standard Claus catalyst. The reaction heats the gases which are cooled to condense elemental sulfur. Part of the cool gas is recycled to the feed in order to control the reaction temperature. The recycled gas dilutes

the H<sub>2</sub>S content to limit the temperature rise. Gas that is not recycled continues to a second and third stage which may use Selectox or less expensive conventional alumina catalyst. After the third reactor the tail gas may be either thermally or catalytically incinerated.

Because the Selectox process is entirely catalytic, no thermal reaction furnace is used, instrumentation and process controls have been greatly simplified resulting in simpler and more reliable operation.

The Selectox catalyst is also in the BSR/Selectox I tail gas cleanup process should the environmental restrictions dictate higher than 96% recovery.

2. CAPITAL INVESTMENT

Capital investment for Recycle Selectox Sulfur Plants should be competitive with an equivalent size three stage Claus Sulfur Plant.

Table 6-1. ANNUAL OPERATING COSTS FOR CLAUS SULFUR RECOVERY PLANTS WITH INCINERATORS<sup>a/</sup>

Component	Plant Size, Mg/d (ltd)		
	5	10	100
Manpower per Shift	1/2	1/2	1/2
Direct Labor @ \$6.41/hr. <sup>b/</sup>	\$ 28,100	\$ 28,100	\$ 28,100
Supv. & Fringes @ 40% DL	11,200	11,200	11,200
Electric Power @ \$.030/kwh <sup>c/</sup>	1,800	3,500	35,300
Fuel @ \$21.20/10 <sup>3</sup> m <sup>3</sup> (\$0.60/MCF) <sup>d/</sup>	2,100	4,100	41,300
Maintenance @ 3% Inv.	22,700	27,200	83,500
Steam Credit @ \$1.10/Mg (\$0.50/MLB) <sup>e/</sup>	(5,400)	(10,800)	(108,400)
Sulfur Credit @ \$24.61/Mg (\$25/LT) <sup>f/</sup>	(41,500)	(83,000)	(830,000)
Net Operating Cost	19,000	(19,700)	(739,000)

a/ 350 operating days per year, 90% H<sub>2</sub>S gas stream.

b/ U.S. Department of Labor, "Monthly Labor Review", October, 1975, p. 96.

c/ Amount from Ford, Bacon, and Davis, "Sulfur Recovery Plants", 1971, page 80; price from "Typical Electric Bills, 1974, FPCR-83", Federal Power Commission.

d/ Process Research, Inc., EPA Contract 68-02-0242, Task 2, 1973, page 75 (LTR. L.L. Zuber, Stauffer Chemical, to W.D. Beers, dtd. 3-21-72).

e/ Ford, Bacon, & Davis, Op. Cit.

f/ Assumes 95% recovery with tail gas treatment.

Table 6-2. ANNUAL COSTS FOR CLAUS SULFUR RECOVERY PLANTS WITH WELLMAN-LORD

Component	Claus Plant Size Mg/d (ltd)		
	5	10	100
Manpower Per Shift <sup>a/</sup>	2	2	2
Direct Labor @ \$6.41/hr. <sup>b/</sup>	\$ 112,400	\$ 112,400	\$112,400
Supv. & Fringes @ 40% DL	45,000	45,000	45,000
Electric Power @ \$0.03/Kwh <sup>b/</sup>	3,800	7,600	75,600
Fuel @ \$21.20/10 <sup>3</sup> m <sup>3</sup> (\$0.50/MCF) <sup>a/</sup>	2,100	4,100	41,300
Maintenance @ 3% INV(Claus)	1,900	26,100	82,000
5% INV(W-L)	18,300	22,500	89,400
Chemicals, @ \$200/ton 50% NaOH <sup>c/</sup>	3,500	7,000	70,000
Soft Water @ \$1.14/m <sup>3</sup> (\$0.30/MGAL) <sup>d/</sup>	100	200	1,700
Steam Credit @ \$1.10/Ma (\$0.50/MLBS) <sup>a/</sup>	(4,800)	(9,600)	(96,400)
Sulfur Credit @ \$24.61/Mgd (\$25/LT) <sup>f/</sup>	(43,700)	(87,500)	(875,000)
Net Operating Cost	\$158,600	\$ 127,800	\$ (454,000)

a/ Trip Report: Standard Oil Company of California El Segundo Refinery, 10/15/73, C.B. Sedman, EPA.

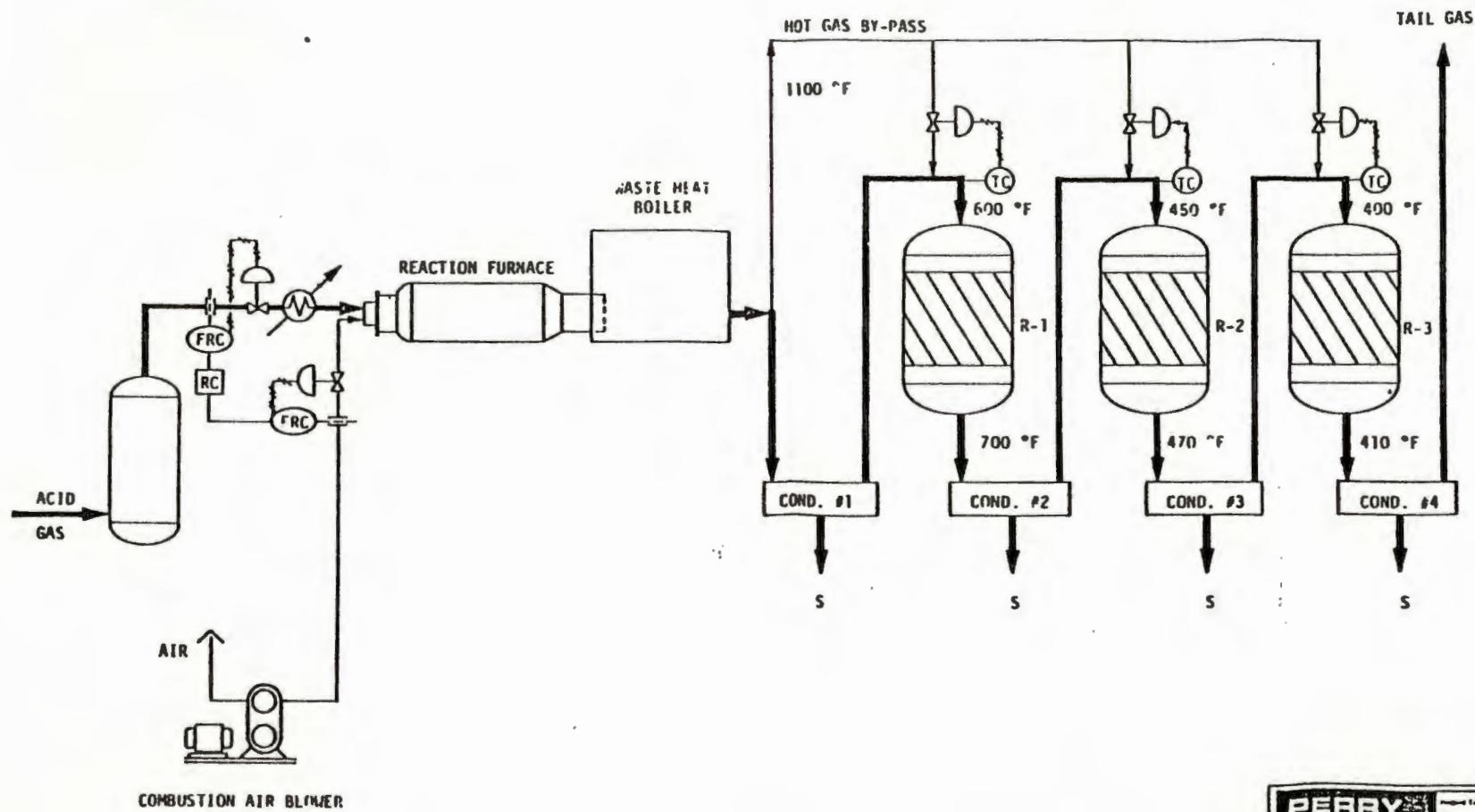
b/ 350 operating days per year.

c/ "Hydrocarbon Processing", April, 1973, p. 116.

d/ Scaling Exponent, 0.3.

e/ Scaling Exponent, 0.6.

f/ Assume 100% sulfur recovery with tail gas unit.



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TELEFAX: 214-422-4400

FIGURE 6-1  
FLOW DIAGRAM

TYPICAL CLAUS PLANT  
FOR SULFUR RECOVERY

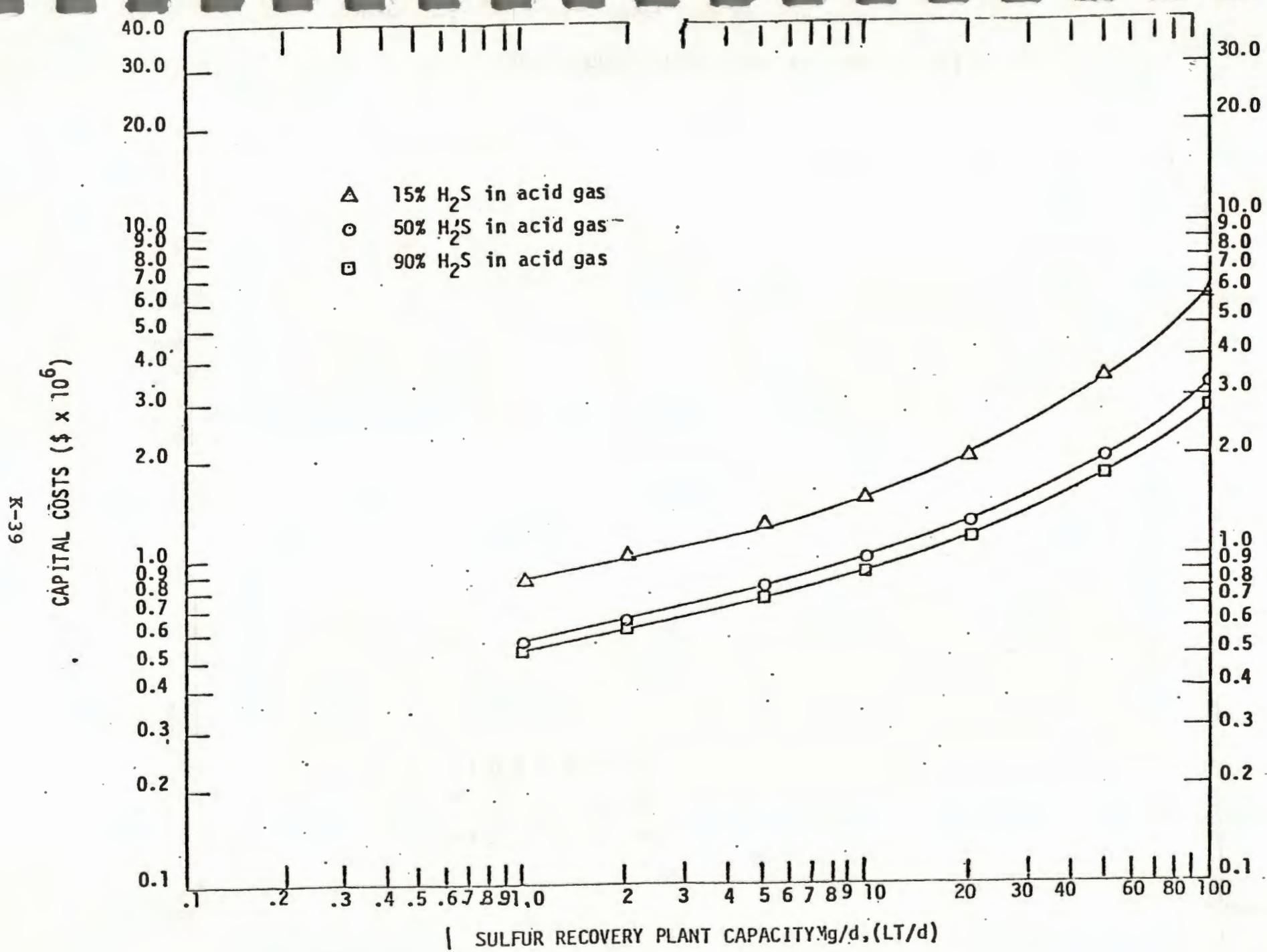


Figure 6-2. Claus (3-stage) plus incinerator - capital cost.

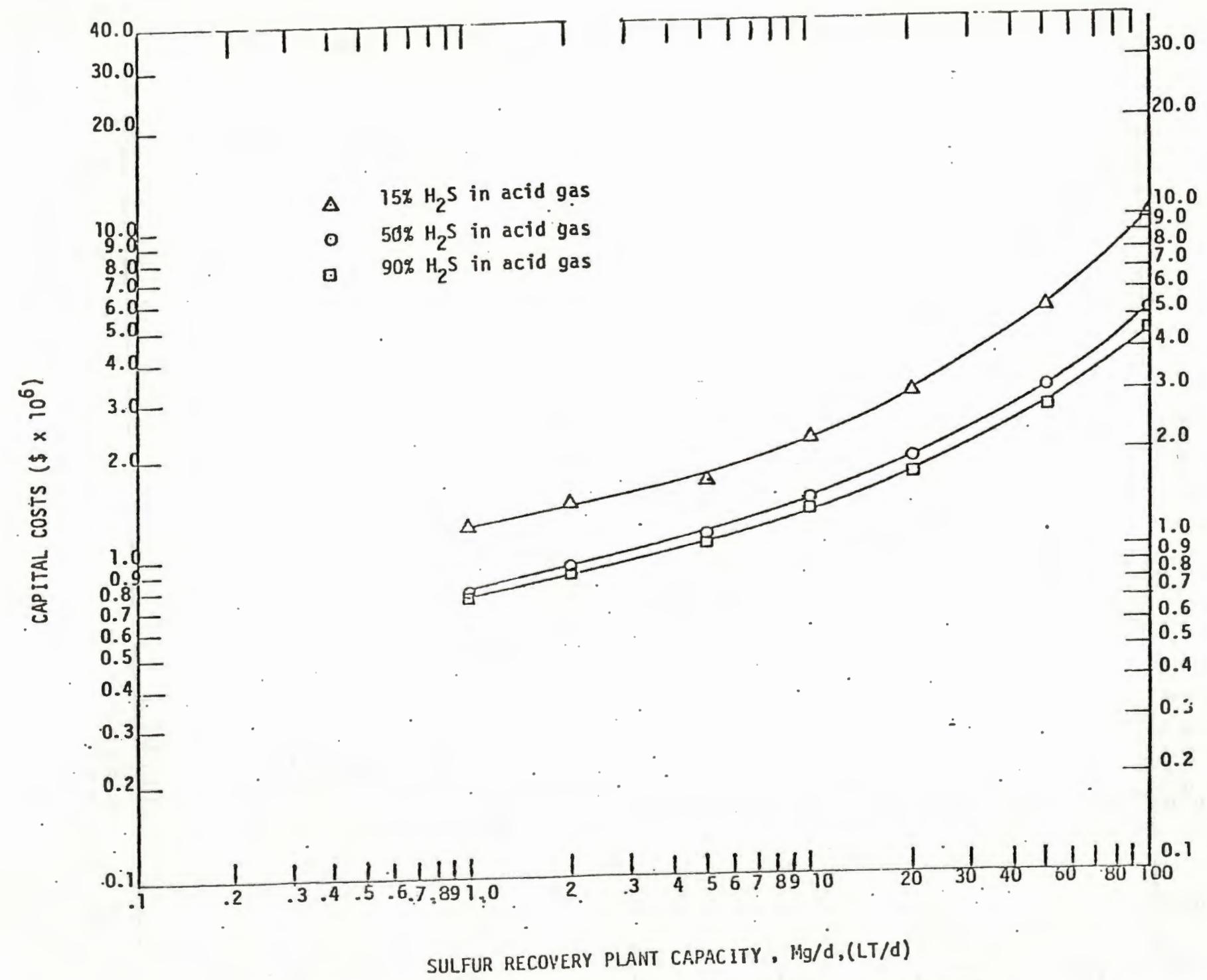


Figure 6-3. Claus (2-stage) plus Wellman-Lord - capital cost.

$$LT/D = (\text{SCFD}/379) * (\% \text{ H}_2\text{S}/100) * (32.06/2240)$$

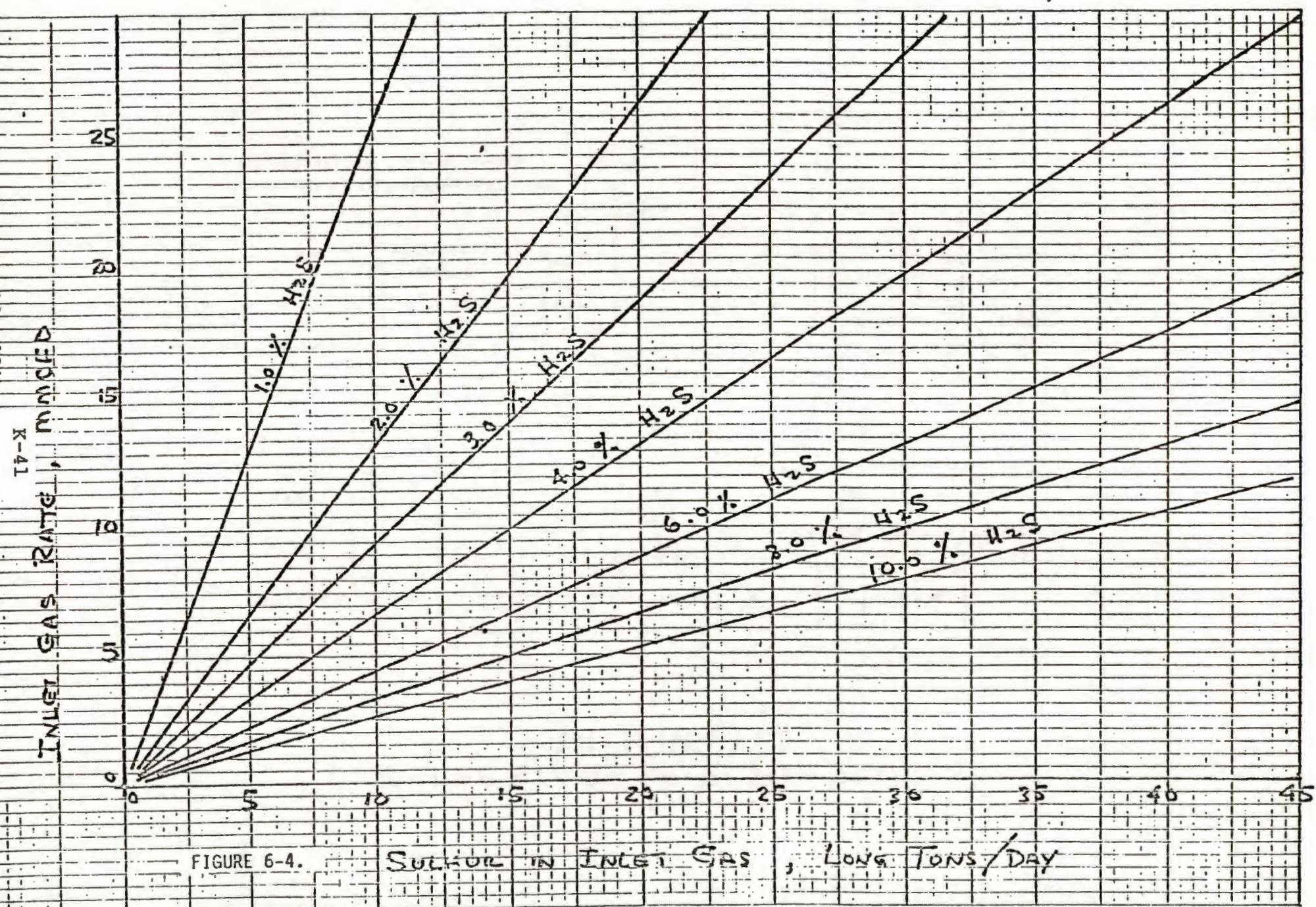
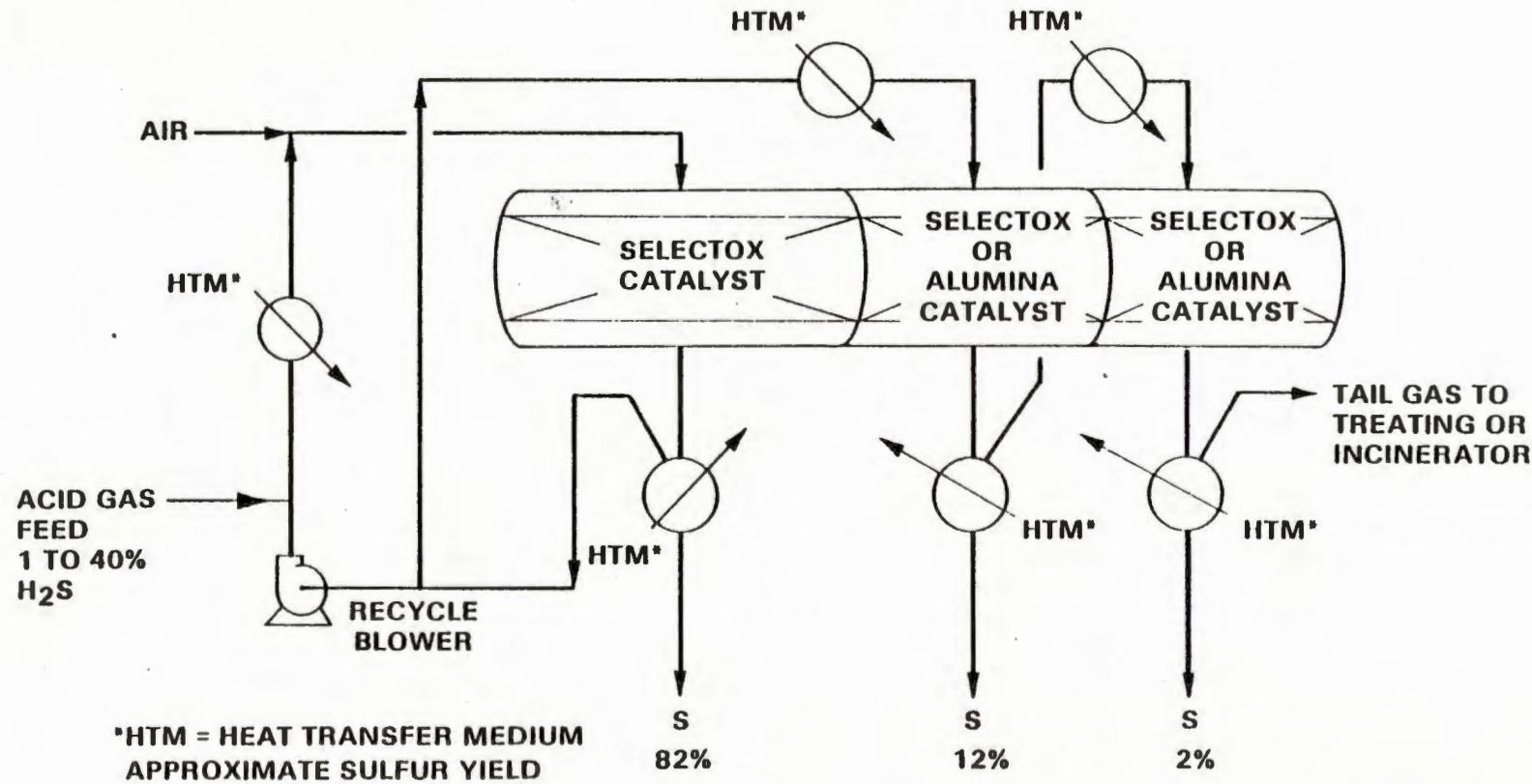


FIGURE 6-4.

SULPHUR IN INLET GAS, LONG TONS/DAY

Figure 6-5

## THREE STAGE SELECTOX PROCESS WITH RECYCLE



## SECTION 7

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