

INTERIM REPORT:

REMOVAL OF H<sub>2</sub>S FROM GEOTHERMAL STEAM  
BY CATALYTIC OXIDATION PROCESS  
BENCH SCALE TESTING RESULTS

C. T. Li  
R. A. Brouns

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Pacific Northwest Laboratory  
Richland, Washington 99352

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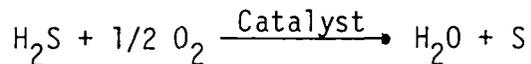
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## SUMMARY AND CONCLUSIONS

Pacific Northwest Laboratory (PNL) has investigated a process to remove hydrogen sulfide (H<sub>2</sub>S) from geothermal steam, under DOE sponsorship. This process is an upstream steam treatment process which utilizes a catalytic oxidation reaction to convert H<sub>2</sub>S in geothermal steam to water vapor and sulfur. The principle process reaction is:



The process consists of passing geothermal steam, containing H<sub>2</sub>S and other noncondensable gases, through fixed beds of activated carbon catalyst. Oxygen is provided by injection of air or oxygen upstream of the catalyst beds. The treated steam, with H<sub>2</sub>S being almost completely removed, passes to steam turbines for power generation. The elemental sulfur produced deposits on the catalyst surface and is retained. The catalyst activity decreases gradually with sulfur accumulation. Sulfur removal, and catalyst regeneration, is accomplished by solvent extraction. Sulfur is recovered from solvent by evaporation/crystallization.

Bench scale experimental work on this process was performed to determine its performance and limits of applicability to power generation systems employing geothermal steam. The bench scale system employed a one-inch diameter reactor, a steam supply with controlled temperature and pressure, an injection system for adding H<sub>2</sub>S and other gases at controlled rates, and instrumentation for control and measurement of temperatures, pressures, flow rates and pressure drop. H<sub>2</sub>S and other analyses were performed by wet chemistry techniques.

On the basis of the bench scale data, the following conclusions were drawn:

- The process removes H<sub>2</sub>S from simulated geothermal steam effectively when the steam to be treated is sufficiently superheated to avoid condensation in the catalyst bed. With an inlet H<sub>2</sub>S concentration of

200 ppm, the treated steam outlet  $H_2S$  concentration will be less than 20 ppm at a catalyst loading of 0.15 lb sulfur/lb carbon, a steam flow rate of 40 lb/min-ft<sup>2</sup>, and steam pressure in the range of 80 to 120 psig. At lower sulfur loading, the  $H_2S$  content of treated steam approaches zero.

- The presence of noncondensable gases such as  $CO_2$ ,  $H_2$ ,  $CH_4$ ,  $NH_3$  and  $N_2$  does not affect the  $H_2S$  removal capability of the oxidation process. Ammonium sulfate may be produced if a large excess of oxygen is present.
- When the temperature of steam being treated is higher than 235°C, the entrainment of sulfur from the carbon bed becomes severe at a steam flow rate of 44 lb/min-ft<sup>2</sup>.
- Oxygen required for effective oxidation of  $H_2S$  ranged between 1.1 to 1.5 times the stoichiometric requirement, depending on the type of activated carbon used as the catalyst.
- The pressure drop of steam across packed catalyst beds can be calculated by a modified correlation for noncondensable gas flow across packed beds.
- A standard industrial grade activated carbon, manufactured from bituminous coal, is the preferred catalyst for employment in the system. No preconditioning of the catalyst is necessary.
- Sulfur can be removed from the carbon catalyst by solvent extraction using  $CS_2$ , tetrachloroethylene, or dichloroethane. Dichloroethane appears to be the preferred solvent owing to its low health hazard potential and inertness in the system.
- Condensation from saturated or wet raw steam in catalyst beds, with subsequent deactivation of the catalyst, can be avoided by adiabatic throttling or heating the raw steam. The latter appears to be more energy efficient.
- Preliminary estimates show a capital investment of \$62.4/kW and an operating cost of 1.27 mil/kWh for a steam treatment system of 50 MWe size. Sulfur credits could reduce the operating cost to 1.15 mil/kWh.

Overall, the catalytic oxidation process for the removal of  $H_2S$  from geothermal steam is feasible and it can effectively remove  $H_2S$  from geothermal steam for reasonable cost. Major potential advantages of the process include: use of a relatively cheap catalyst; use of air as an oxidant; direct recovery of elemental sulfur; and little or no by-product and process waste disposal problems.

It is believed that the proposed process offers sufficient promise to warrant investment in the next phase of development and evaluation. This would consist of engineering scale evaluation at one or more test sites where geothermal steam is available.

This engineering scale evaluation would involve the design, construction and operation of a movable (skid mounted), self-contained test unit. It would include all necessary features of an integrated process, including sulfur recovery and superheating the raw steam, if dictated by anticipated raw steam quality.

The objectives of this pilot scale work should include:

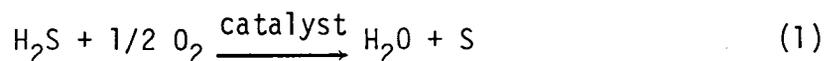
- Establish  $H_2S$  removal efficiency and capacity with real geothermal steam--particularly with contaminants which could not be simulated in the laboratory (such as mercury and borates).
- Establish catalyst life with a real geothermal steam source.
- Determine the effects of the process on steam corrosivity.
- Develop the design features necessary for integration with geothermal power systems (solids removal, steam superheating, materials selection, instrumentation and controls).

## INTRODUCTION

Geothermal fluids from most sources contain hydrogen sulfide ( $H_2S$ ). The use of these energy containing fluids for power generation, space heating and other industrial applications is hampered because of the presence of this noxious gas.  $H_2S$  is corrosive to power generation-transmission equipment as well as offensive to the environment. Shut-down of geothermal power plants due to corrosion failures related to the presence of  $H_2S$  and the close relation between power plant operating factor and the  $H_2S$  concentration in geothermal steam have been reported.<sup>(1)</sup> The release of  $H_2S$  containing effluents from both geothermal power plants and well drilling sites will become an increasingly severe problem as more stringent pollution control and environmental protection regulations are implemented. Practical, economical, and efficient methods for avoiding these problems are needed.

The objective of the work described in this report was to develop an effective process which will solve both corrosion and environmental problems encountered in the utilization of geothermal steam. The process concept is based on the catalytic oxidation of  $H_2S$  to elemental sulfur. It is intended to be employed upstream of power generation equipment, as shown in Figure 1, so as to mitigate both equipment corrosion and  $H_2S$  emission problems.

The chemistry of the proposed oxidation process can be represented by Equation (1).



The catalyst is activated carbon.

In the overall process concept, steam from geothermal wells passes through reactors packed with activated carbon catalyst. In the presence of the catalyst  $H_2S$  reacts with oxygen according to Reaction (1). Oxygen is introduced to the reaction system either as pure oxygen or air. Steam, after being treated, leaves the reactors for power generation. Sulfur produced from the reaction deposits on the surface of the carbon and remains in the reactor. The exhaust steam from

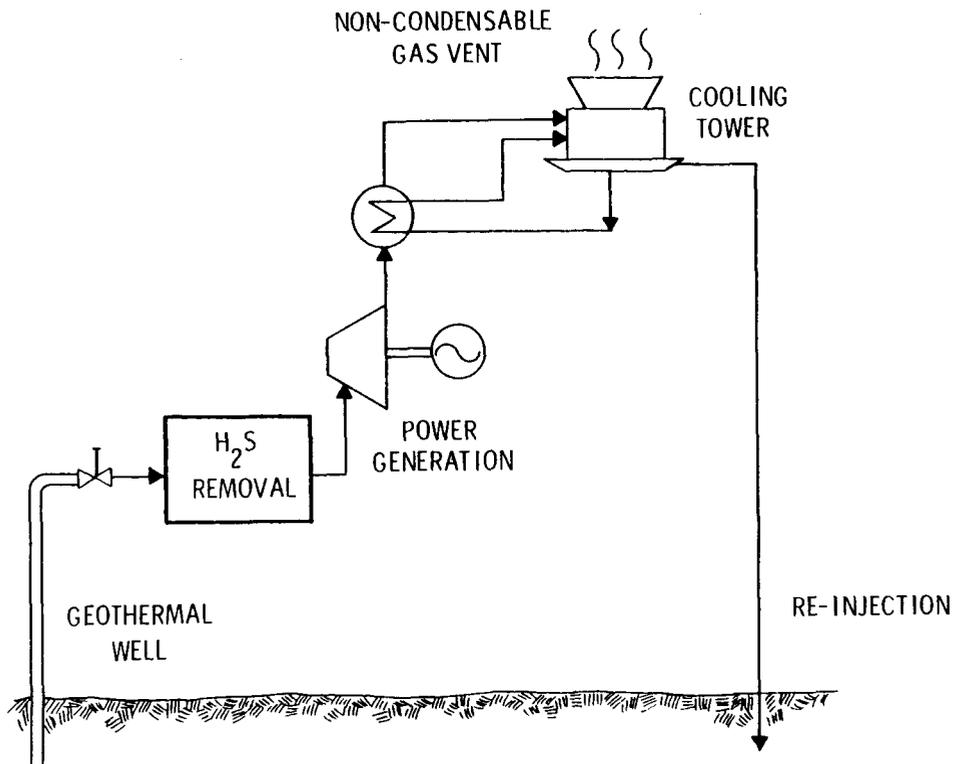


FIGURE 1. Up-Stream Geothermal Steam Treatment Process

the turbine power-generator is condensed and then cooled in a cooling tower. Water from the cooling tower is finally injected into re-injection wells.

The activity of the catalyst will decrease as sulfur accumulates on its surfaces. Regeneration of spent catalyst can be accomplished by a solvent extraction process. Elemental sulfur is recovered from the solvent and the regenerated carbon, with or without a catalyst re-activation process, is ready for reuse. A simplified block diagram of the proposed catalytic oxidation process for the removal of  $H_2S$  from geothermal steam is shown in Figure 2.

The proposed process evolved from previous PNL work on the use of solid sorbents which would act as acceptors for  $H_2S$  from geothermal steam. It was found that a few solid sorbents, for example  $ZnO$  and a synthetic resin, were able to remove  $H_2S$  from geothermal steam. However, the sorbent regeneration was found to be either technically infeasible or uneconomical. Exploratory experiments during the course of that investigation showed that  $H_2S$  could be catalytically oxidized to form sulfur and water. Removal of  $H_2S$  from gas streams

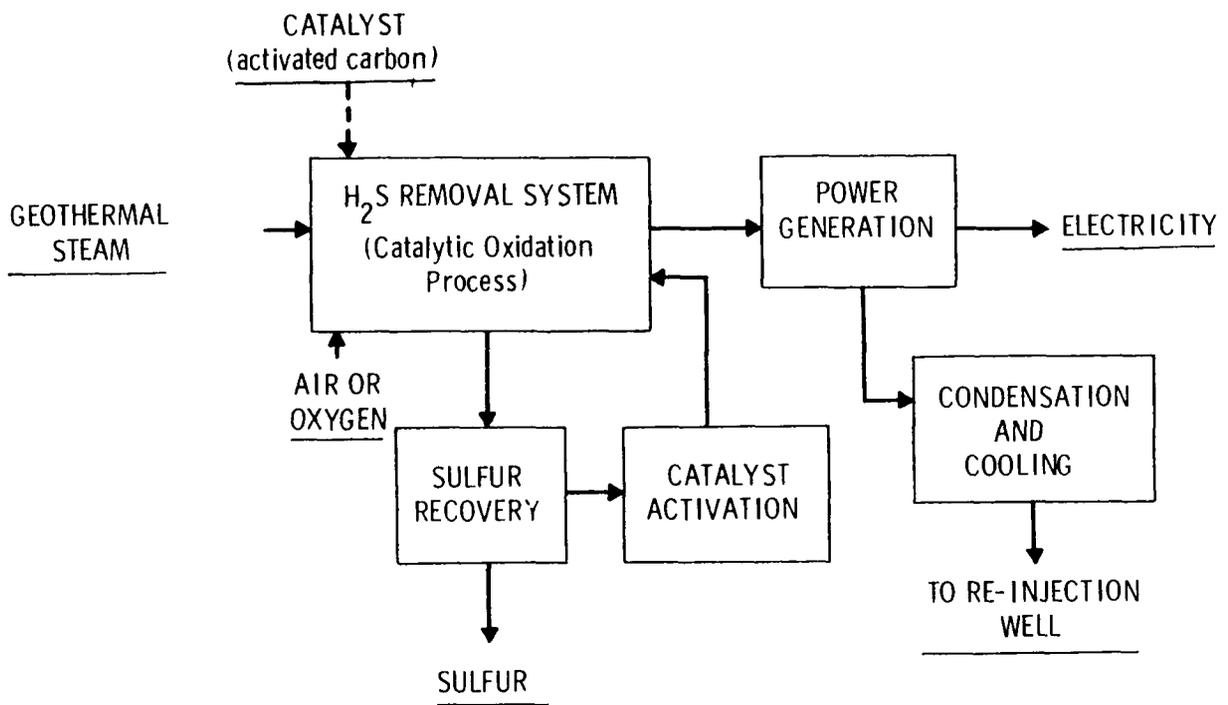


FIGURE 2. General Flow Diagram of the Proposed Catalytic Oxidation Process for  $H_2S$  Removal from Geothermal Steam

by the catalytic oxidation process is an old technique;<sup>(2-6)</sup> however, the application of this old technique to the treatment of steam is new.



## DISCUSSION

### THERMODYNAMICS AND CHEMISTRY OF H<sub>2</sub>S OXIDATION

The use of catalytic oxidation processes for the removal of H<sub>2</sub>S from various gases, primarily fuel gases, has been widely practiced in the past.<sup>(2-6)</sup> However, since water is a coproduct of the oxidation ( $\text{H}_2\text{S} + 1/2 \text{O}_2 \rightarrow \text{H}_2\text{O} + \text{S}$ ), the potential efficiency of such a process in a pressurized steam atmosphere must be considered. Using standard values, the heat of reaction at 25°C was calculated to be -52.98 Kcal/g-mole, indicating that the reaction is highly exothermic. The free energy of the reaction ( $\Delta G$ ) is -46.74 Kcal/g-mole at 25°C. The equilibrium constant may be calculated from the relation

$$\Delta G = -RT \ln K \quad (2)$$

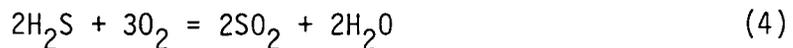
as  $1.92 \times 10^{34}$  at 25°C and  $1.87 \times 10^{21}$  at 175°C. The assumptions that concentrations are represented by partial pressures, that all activities are equal to unity, and that the ideal gas law prevails results in the following expression:

$$K = K_p = \frac{P_{\text{H}_2\text{O}}}{(P_{\text{H}_2\text{S}})(P_{\text{O}_2})^{1/2}} \quad (3)$$

At representative pressures for geothermal steam and using the stoichiometric amount of oxygen necessary for complete H<sub>2</sub>S oxidation, the equilibrium partial pressure of H<sub>2</sub>S would nearly disappear. Based on equilibrium considerations alone, it is concluded that the conversion of H<sub>2</sub>S to S would be essentially complete at representative geothermal steam conditions.

In spite of favorable equilibrium considerations, it is well known that the rate of reaction in gas phase H<sub>2</sub>S oxidation at modest temperatures is too slow for practical applications. Catalysts are therefore employed.

Besides the main reaction, side-reactions<sup>(7)</sup> may also take place in the steam-H<sub>2</sub>S-O<sub>2</sub>-activated carbon system. The possible side reactions are:



SO<sub>2</sub> formed in Reaction (4) can be further oxidized to SO<sub>3</sub> according to the following reaction:

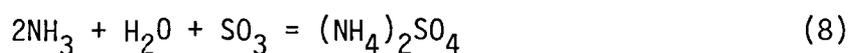


In the presence of metal oxide SO<sub>3</sub> may be converted to MSO<sub>4</sub> as shown in Reaction (7).



Where MO and MSO<sub>4</sub> represent metal oxides and metal sulfates, respectively.

Ammonia, one constituent of the noncondensable gases in geothermal steam can also react with SO<sub>3</sub> to form (NH<sub>4</sub>)<sub>2</sub>SO<sub>4</sub> according to Reaction (8).



The presence of MSO<sub>4</sub>, for example CuSO<sub>4</sub> and (NH<sub>4</sub>)<sub>2</sub>SO<sub>4</sub>, has been confirmed in our experiments.

## BASES AND DEFINITIONS

In order to standardize all laboratory work and to make discussions easier in this report, the following bases will be followed:

### Composition and Conditions of Geothermal Steam

Although experimental results are broadly applicable to a spectrum of geothermal fluids, steam conditions (temperature and pressure) and compositions used in the work were modeled after those at The Geysers, California. The typical composition and conditions of this steam are shown in Table 1. Simulated geothermal steam, which was prepared by mixing steam generated from a boiler with noncondensable gases such as CO<sub>2</sub>, H<sub>2</sub>, H<sub>2</sub>S, CH<sub>4</sub>, NH<sub>3</sub> and N<sub>2</sub>, was used in all experiments unless otherwise mentioned specially. Although the simulated steam nominally corresponded to Table 1 values, some variations were unavoidable due to controller limitations.

TABLE 1. Typical Chemical Composition of Geothermal Steam<sup>(9)</sup>  
Produced in The Geysers

Wellhead Temperature, °F	347
Total Noncondensable Gas Content of Steam (wt%)	0.7
Noncondensable Gases in Steam (ppm by weight)	
CO <sub>2</sub>	6,210
H <sub>2</sub>	52
H <sub>2</sub> S	208
CH <sub>4</sub>	385
NH <sub>3</sub>	55
N <sub>2</sub>	91

### Breakthrough Time

The breakthrough time was defined as the time required for the outlet H<sub>2</sub>S concentration to reach ten percent of that of the inlet. For example, if the inlet H<sub>2</sub>S concentration is 200 ppm, the breakthrough time will be the time when the outlet H<sub>2</sub>S concentration reaches 20 ppm.

### Catalysts

Catalysts used in this study were mainly activated carbon. Various brands of carbon were obtained or purchased from different manufacturers or distributors including Barnebey Cheney, Fisher Science Co., Calgon Corporation and Girdler Chemical, Inc. Carbons used varied in shape and size. Some were impregnated with metals or metal oxides.

Information from the manufacturers is given in Table 2 regarding applications, properties, and, in some cases, manufacturing method.

### Size of Power Plant

In this study the size of a typical geothermal power plant was taken as 50 MWe using 1,000,000 lb/hr of steam.

**TABLE 2. Manufacturer's Recommended Application and Physical Properties of Activated Carbons Used in Bench Scale Experiments**

Sample	Manufacturers' Recommended Applications	Physical Properties
1	For recovery of gasoline, benzol, ether, alcohol, light oils and other solvents from natural gas, still gases, air, vapor lines, and similar gases.	6 to 14 mesh granules. It is impregnated with a metal compound.
2	Specially treated activated carbon formulated for the removal of sulfur compounds, including hydrogen sulfide, carbonyl sulfide and mercaptans, from hydrocarbon streams in fix-bed vapor phase operation.	Prepared by chemical impregnation of a physically strong, highly absorbent activated coconut shell char; containing approximately 5 weight percent copper, chloride free non-corrosive material. Surface Area: 900 sq cm/g, Bulk Density: 35 to 38 lb/cu ft, Pore Volume: 0.6 cc/g. Color: black, Particle size: 4 x 8 mesh granules
3	For removal of hydrogen sulfide and methyl mercaptan typically found in sewage treatment, pulp and paper mills, petroleum refineries and chemical plants.	Bulk Density: 34 lb/ft <sup>3</sup> , Particle Size: 4 x 10 mesh, Hardness Number, Minimum: 90, Moisture, Maximum: 15%, Voids in Dense Packed Column: 43%
4	For vapor phase application; solvent recovery system, recovery of alcohols, chlorinated hydrocarbons, ethers, ketones, esters, hydrocarbon, and aromatics; used as the catalyst support in the acetylene process for the production of vinyl chloride and vinyl acetate monomers, also used for the separation of hydrocarbon gas streams, such as the recovery of C <sub>3</sub> and C <sub>4</sub> cuts from natural gas; removal of organic sulfur, COS, and higher hydrocarbons from methane and hydrogen for catalytic conversion processes; purification of carbon dioxide for beverage use and dry ice; removal of chlorine, chlorinated organics and aromatics from anhydrous hydrogen chloride; purification of acetylene, hydrogen, compressed air, etc.	Particle Size: 4 x 10 mesh, Surface Area: 1050-1150 m <sup>2</sup> /g, Bulk Density: 30 lb/ft <sup>3</sup> , Particle Density: 0.85 g/cc, Real Density: 2.1 g/cc, Pore Volume: 0.7 cc/g, Voids in Dense Packed Column: 43%, Specific Heat at 100°C: 0.25, Iodine Number, Minimum: 1050, Carbon Tetrachloride Adsorption, Min. Wt: 60%, Ash, Maximum: 2%, Hardness Number, Minimum: 90-93
5	For removal of hydrogen sulfide and low molecular weight, organic sulfur compounds from gas streams, odor control, product purification and elimination of those sulfur compounds that affect catalyst performance	Made from bituminous coal and suitable binders specially impregnated, granular product. Particle Size: 12 x 30 mesh, Bulk Density: 32 ~ 35 lb/ft <sup>3</sup>

TABLE 2. Manufacturer's Recommended Application and Physical Properties of Activated Carbons Used in Bench Scale Experiments

<u>Sample</u>	<u>Manufacturers' Recommended Applications</u>	<u>Physical Properties</u>
6	For water treatment, removal of odor from drinking water	12 x 30 mesh granules
7	For removal of H <sub>2</sub> S, mercaptans, and other organic sulfur compounds (especially from natural gas)	6 x 10 mesh granules impregnated with chemicals to enhance its ability to adsorb and retain specific gases
8	Removal of hydrogen sulfide and hydrogen cyanide from air, inert gases and other non-oxidizing atmospheres	Made from coconut shell and activated with high temperature steam. Adsorption characteristics: 60 minute (typical) U.S. Government accelerated chloropicrin test (modified) to breakthrough point on sample ground to 6 x 14 mesh (MIL-C-17605-B), 60% (typical) carbon tetrachloride by standard test to saturation (MIL-C-17605-B), Nominal mesh size (6 x 10 Tyler Standard Screen). Hardness: greater than 95% U.S. Government ball abrasion test (MIL-C-17605-B), Bulk Density: 0.50 gram/ml, dense packing. Ash: 12-16%, Moisture content: 5% as packed.
9	Removal of medium and high concentrations of organic vapors from air. Purification of gases. Solvent recovery, Gas separation, high capacity oil vapor removal. Catalyst support.	Made from coconut shell, and activated with high temperature steam. Adsorption Characteristics: 75 minutes (ground 6 x 14), Nominal mesh size: 6 x 10 (Tyler Standard Screens); Hardness: greater than 95% as measured by ball abrasion test; Bulk density: 0.43 to 0.49 grams per ml (dense packing) 27 to 30 lb per cubic foot; Ash: 5% (typical); moisture content: 5% maximum, as packed.

## EXPERIMENTAL RESULTS

### Experimental Bench Scale System

All  $H_2S$  removal experiments were carried out in a bench scale system. This system consists of a boiler feed water pump, steam generator, steam regulator, steam condenser, gas-liquid separator and a noncondensable gas mixing station. The instruments monitoring various operating parameters include a wet testmeter, a temperature recorder, thermocouples, gas flow meters and a differential pressure transducer. All parts contacting the simulated geothermal steam are made of stainless steel. Details of the bench scale system are given in Appendix I.

To evaluate  $H_2S$  removal effectiveness, catalyst bed inlet and exit  $H_2S$  concentrations must be monitored. Hewlett Packard research gas chromatographs with columns of different packing materials and a flame photometric sensor were initially utilized. However, it was soon found that reproducibility was poor and this technique was abandoned. The poor reproducibility was mainly due to the large amount of water vapor in the samples which caused a tailing problem in the chromatogram. Also, condensation of steam in the sampling lines made it impossible to obtain consistent samples. Difficulties in using the gas chromatographic technique in analyzing gas samples in a geothermal steam system have been reported by others.<sup>(8)</sup>

A wet chemical analysis technique, based on an ASTM method was subsequently developed. Basically an iodometric back-titration method is employed to determine the concentration of  $H_2S$  which is captured in an  $H_2S$  fixing reagent such as zinc acetate or cadmium acetate aqueous solution. Details of the  $H_2S$  analysis procedure including a sampling procedure are given in Appendix I.

In all, a total of 77 experiments were performed with the bench scale system. Typically, each experiment consisted of measurement of  $H_2S$  removal effectiveness and time to breakthrough under specific conditions of steam composition, flow rate, temperature, and pressure and with a specific catalyst and catalyst condition. Each experimental run required about 12 hours. Summary data sheets are reproduced in Appendix II for each run discussed in this report.

## Evaluation of Catalyst Performance

The performance of each of the nine kinds of activated carbon listed in Table 2 has been evaluated. In each evaluation experiment 80 grams of carbon was charged into the one-inch diameter reactor. Depending upon the kind of carbon charged, the catalyst bed height varied from 9 to 15 inches. Superheated steam (100 psig, 180 to 185°C) at a flow rate of 44 lb/ft<sup>2</sup>/min (22 g/cm<sup>2</sup>/min) and H<sub>2</sub>S concentration of 200 ppm (nominal) was fed to the reactor. The space velocity under these conditions ranged from 130 to 200 (volume/(volume))/min. The amount of air mixed with the steam was adjusted so that the oxygen introduced to the reaction system was about 1.5 times the stoichiometric requirement of the oxidation reaction, Reaction (1).

Gas samples of inlet and outlet streams of the reactor were taken and H<sub>2</sub>S concentration was determined by the gas analysis method described in the previous section. A catalyst performance evaluation test would last from 8 to 13 hours, depending on when breakthrough occurred. When each experiment was completed, the spent carbon in the reactor was discharged and dried in a vacuum oven (100°C). Sulfur, which deposited on the surface of the spent carbon was recovered by solvent extraction. The extraction of sulfur was carried out in a Soxhlet using CS<sub>2</sub> as a solvent.

Among these tests the shortest breakthrough time (as defined previously) was less than three hours and the longest was 27.8 hours. The pertinent data show obtained from these experiments are summarized in Table 3. These data show that activated carbon Samples 1 and 5 most effectively removed H<sub>2</sub>S from the H<sub>2</sub>S-steam mixture. Breakthrough times for Sample 1 and Sample 5 are 27.8 and 19.2 hours, respectively, and overall H<sub>2</sub>S removal for them is higher than 97.4 percent. Samples 1 and 5 are laboratory and industrial grade materials, respectively; therefore, carbon Sample 5 was chosen and used in most of the subsequent experiments, except as noted.

TABLE 3. Summary Data for Catalyst Evaluation Tests

Run No.	163	117	119	123	124	125	115	127	128
Catalyst	1	2	3	4	5	6	7	8	9
Wt. of Cat. gm	80	80	80	80	80	80	80	80	80
Bed Volume, cm <sup>3</sup>	170.7	136.9	128.1	157.5	128.8	184.0	114.1	143.5	164.9
Steam Rate, gm/cm <sup>2</sup> /min	22.66	22.42	22.00	19.8	20.2	20.5	20.5	20.6	20.8
Space Velocity, /min	156.7	192.4	201.9	148.0	184.6	130.9	211.1	168.9	148.4
Residence time, sec	0.38	0.31	0.30	0.41	0.33	0.46	0.28	0.36	0.40
Air Rate, ml/min	50	50	50	50	50	50	50	50	50
Oxygen Stoichiometric Ratio	1.44	1.48	1.43	1.78	1.59	1.52	1.57	1.71	1.49
Inlet H <sub>2</sub> S Conc., ppm	193.0	191.2	202.3	191.6	197.9	205.2	199.4	181.8	207.1
1 hr	0.58	7.2	1.2	3.8	0.5	0.6	1.0	1.3	1.5
2	0.68	15.7	1.5	7.7	1.0	1.1	1.2	2.0	2.2
3	0.77	23.5	2.0	12.8	1.3	1.5	1.8	2.8	3.3
4	0.89	32.0	5.0	17.3	1.8	1.8	2.0	3.9	4.3
5	1.00	39.7	12.3	22.1	2.2	2.0	2.6	4.9	7.5
6	1.15	47.8	20.0	26.3	2.5	2.3	3.2	6.1	12.4
7	1.30	55.5	27.3	30.8	3.1	5.2	3.9	6.8	20.3
8	1.50	63.6	34.8	34.7	3.8	8.2	4.6	7.5	20.3
9	1.75			38.7	4.0	12.7	6.8	8.3	24.2
10	1.97			43.2	4.8	16.5	9.8		28.1
11	2.24				5.2				
12	2.55				5.8				
13	2.91				6.3				
Breakthrough Time, hr	27.8	2.7	6.2	4.7	19.2*	9.8*	13.0*	10.7*	8.5
Average Outlet H <sub>2</sub> S, ppm	5.9	34.9	17.4	23.0	3.3	4.0	3.0	4.4	11.8
Outlet Oxygen Conc. ppm	42.5	59.4	49.8	72.4	57.5	52.4	54.7	62.2	52.9
Overall H <sub>2</sub> S Removal, %	97.42	81.7	91.4	82.8	98.4	98.1	98.5	97.5	94.3

\*Estimated Values

## Temperature and Pressure Limitations

The oxidation of  $\text{H}_2\text{S}$  to elemental sulfur is exothermic. Chemical thermodynamics would qualitatively predict increasing conversion to sulfur with decreasing temperature. However, from the reaction kinetics point of view, increasing temperatures will generally increase reaction rates. The acceptability of both the rate of reaction and the conversion at geothermal steam temperatures is a key issue to the ultimate feasibility of the proposed process.

The melting point of sulfur is 112 or  $120^\circ\text{C}$  depending upon the type of sulfur formed. Sulfur deposited on the surface of the catalyst may be entrained by steam passing through the catalyst bed if the operating temperature is above the melting point of sulfur. Entrained sulfur may deposit on the surface of downstream turbine blades, casing or exhaust lines depending where the cold spots are.

In Reaction (1) the number of moles of gaseous product is less than that of reactants. According to Le Chatelier's principle<sup>(11)</sup> an increase in the reaction system pressure will enhance the equilibrium yield of the reaction. The effect of pressure on the conversion of  $\text{H}_2\text{S}$  to sulfur is therefore of interest.

To investigate the effect of temperature and pressure on the proposed process carbon sample No. 1, a standard grade of coconut shell activated carbon, was selected as the reference catalyst. All experiments were performed at a standard steam flow rate ( $44 \text{ lb/hr-ft}^2$ , nominal), and a standard  $\text{H}_2\text{S}$  inlet concentration (200 ppm nominal). The same carbon sample was used for all experiments in this series. Catalyst regenerating, after each run, was performed by extracting the adsorbed sulfur with  $\text{CS}_2$  and vacuum drying.

Data from runs 93, 94, 98, 99, 103, 105, 108 and 109, Appendix II, are shown plotted on Figure 3. Although the data are scattered, the data for 110 psig show a lower effectiveness. However, this is attributed to aging of the catalyst, which is known to have contained metal compounds. Since the 110 psig runs were the last to be performed during this series, it is likely that

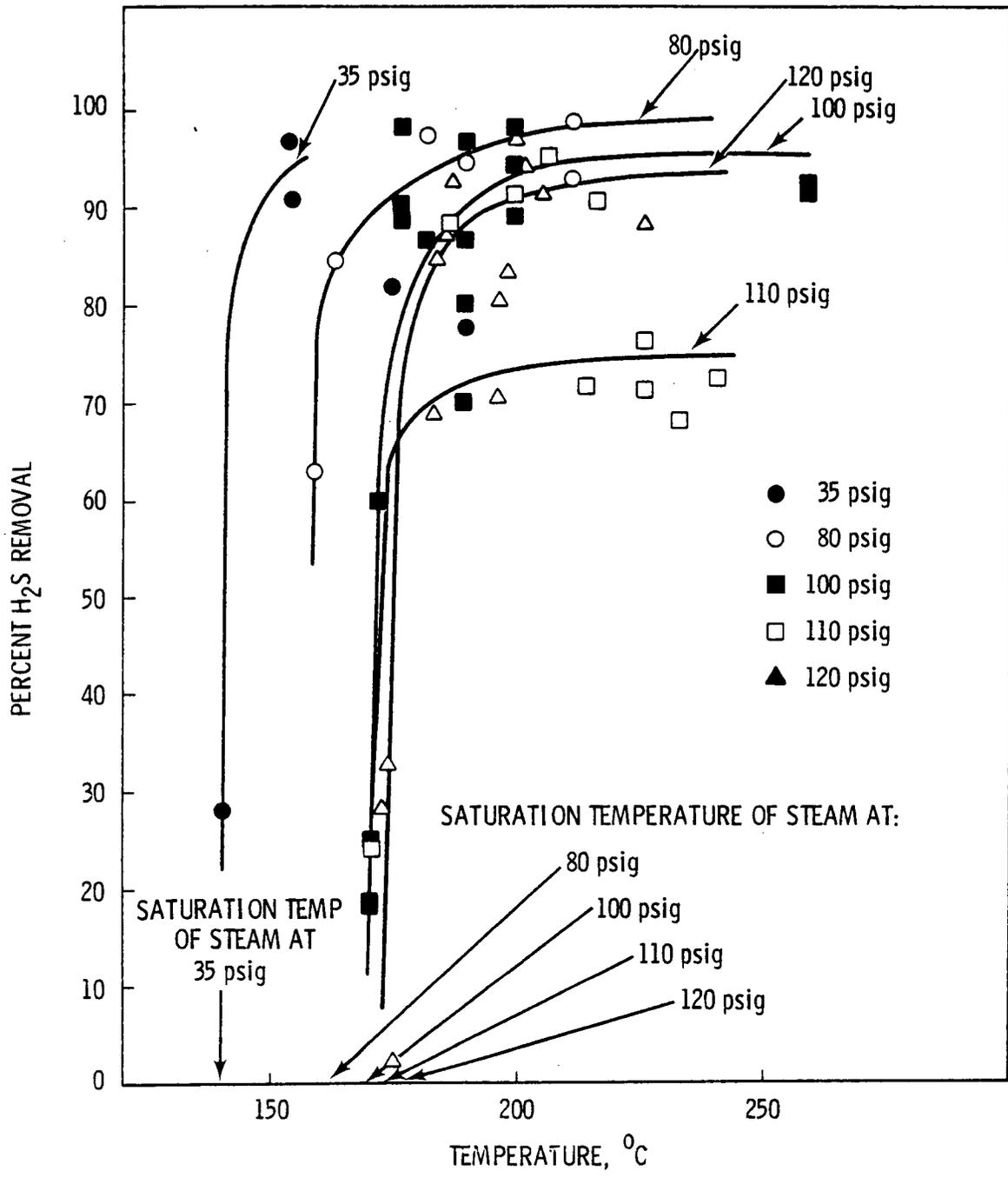


FIGURE 3. Effect of Temperature and Pressure on H<sub>2</sub>S Removed Effectiveness

insoluble metal sulfides formed on the catalyst surface to reduce the activity of the catalyst.

It is evident that  $H_2S$  removal effectiveness rapidly diminished as the steam temperature and pressure approached saturation. The scattering of data can be attributed to:

- (1) These experiments were carried out during the early part of research program. Control of flow rates and temperature had not been resolved at that time.
- (2) Sampling techniques had not been fully developed.

Reasons for the poor performance of the process at the steam saturation point were thought to be due to blinding the active sites of the catalyst by steam condensate. A detailed investigation of this phenomenon will be discussed in the next section.

Overall, it is concluded that neither temperature nor pressure have an intrinsic effect on  $H_2S$  removal effectiveness over the ranges of 140 to 240<sup>0</sup>C and 35 to 120 psig. However, again, the catalyst does become ineffective as the pressure-temperature condition of the steam approaches saturation.

When the steam temperature was higher than 235<sup>0</sup>C, the entrainment of sulfur by steam flowing through the reactor became significant. The entrainment of sulfur was detected by the appearance of a white milky sulfur precipitate in the steam condensate. The entrainment of sulfur was severe, especially when the activated carbon was loaded with sulfur. To determine the amount of sulfur entrained versus steam temperature, a bed of heavily sulfur-loaded spent carbon was purged with steam at different temperatures but at a constant pressure and velocity. The temperature of the steam was increased gradually. The steam condensate was collected and a known amount of condensate was dried in a Petri dish. The residual solid was weighed to determine the solid content. As shown in Run 96 of Appendix II and Figure 4, the residual solid, which was taken as the amount of sulfur entrained, increased as the temperature of the steam increased. Note that no

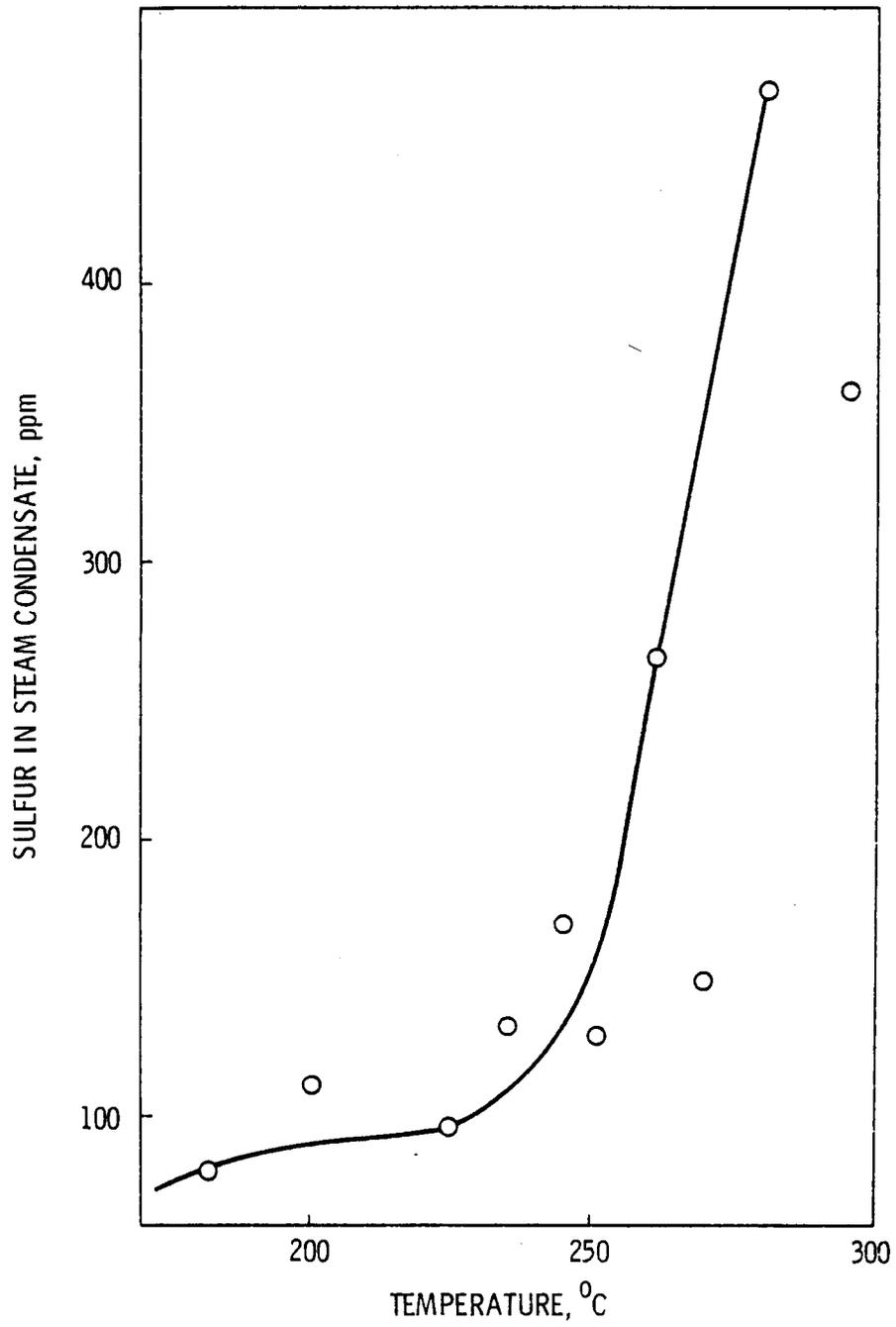


FIGURE 4. Effect of Temperature on Sulfur Entrainment

quantitative deductions should be made from this data due to the possible deposition of sulfur in the connecting piping.

### Pressure Drop

Degradation of steam quality in a geothermal power generation system, by frictional steam pressure drop, would reduce the electrical power output of the system. Since the proposed catalytic oxidation system is intended to be used upstream of power generation equipment, this factor must be considered in evaluating the feasibility of the system.

Pressure drop measurements were made using the bench scale system. These measurements were made during experiments involving  $H_2S$  oxidation as well as in experiments specifically designed for pressure drop measurement with clean steam flow across clean activated carbon beds. The instrument used for the measurements was a strain gauge differential pressure transducer. It was connected to the bench scale reactor just above and below the position of the carbon bed to minimize velocity effects which could occur if the pressure taps were in the connecting piping. Laboratory calibration of the  $\Delta P$  instrument was by means of a mercury manometer.

Initially, the pressure drop experiments were performed with the bench scale reactor loaded with glass beads of the approximate size range of the activated carbon to be employed later. This was done to check the instrument system and to provide a comparison of pressure drop across beds having well characterized media with the pressure drop through beds having a range of particle sizes and sphericity. Typical data for glass beads are shown plotted on Figure 5. There is considerable scatter in the data points obtained with steam at saturation while steam superheated about  $10^{\circ}C$  shows a more consistent relationship between pressure drop and flow. This difference is attributed to partially filling flow passage voids with condensate from saturated steam. Since this is a rather unpredictable occurrence, it is not possible to accurately predict the steam pressure drop across beds in which condensation may occur. It appears that such condensation would increase the bed pressure drop by 40-60% in the velocity range of interest.

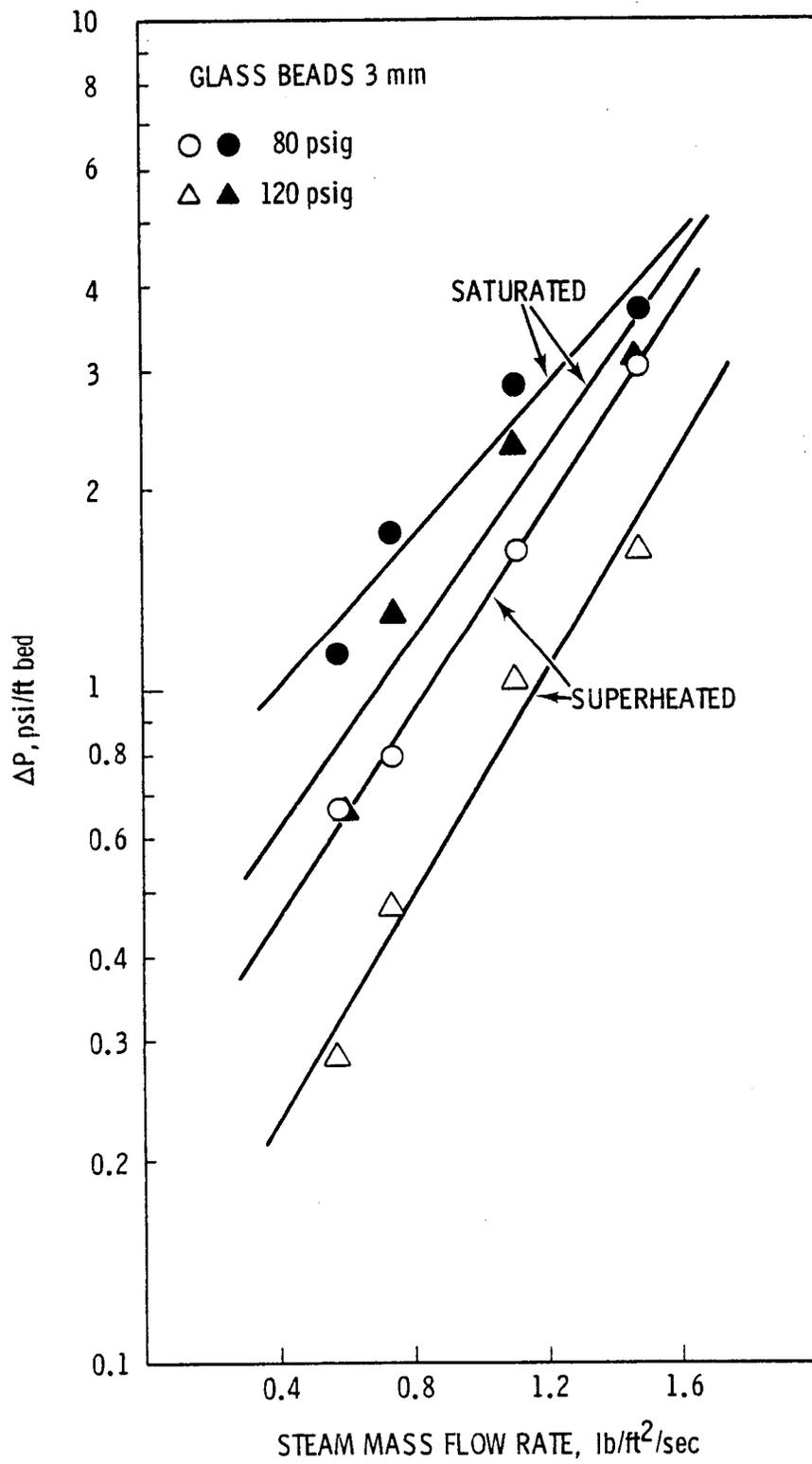


FIGURE 5. Steam Flow Rate vs Pressure Drop at Different Pressures for 3 mm Glass Beads

Experiments to measure pressure drop across carbon beds were performed using clean steam (no additives) and clean carbon beds. Various steam qualities and flow rates were examined. Five levels of steam temperature--162, 166, 170, 177 and 180°C; three pressures--80, 100 and 120 psig corresponding to saturated temperatures of 162, 170 and 177°C respectively were used. Steam flow rates of 17.28, 21.68, 32.52 and 43.06 lb/ft<sup>2</sup>-min were examined. Three kinds of activated carbon, samples 2, 5 and 6 were used. Size and bed properties for these materials are shown in Table 4.

TABLE 4. Activated Carbon Bed Properties

<u>Nominal Size Range</u>	<u>Average Particle Size, mm</u>	<u>Bed Void Volume, %</u>
Sample 2 - 4 to + 8 mesh/in.	2.016	45.6
Sample 5 - 12 to +30 mesh/in.	0.692	41.3
Sample 6 - 12 to +30 mesh/in.	0.526	42.8

Average particle sizes were determined by screen analysis followed by weighted averaging based on particle surface area assuming spherical particles. Six cuts for each nominal particle size were obtained. Bed void volume was determined by prewetting samples with kerosene, to fill internal carbon pores, followed by bulk kerosene displacement measurements.

The pressure drop across catalyst beds is a function of following variables: pressure, temperature, flow rate and particle size. These values were plotted in the form of Ergun's<sup>(12)</sup> correlation as shown in Equation 9 and figures 6 and 7.

$$\left(\frac{\Delta P \bar{\rho} g_c}{G_o^2}\right) \left(\frac{D_p}{L}\right) \left(\frac{\epsilon^3}{1-\epsilon}\right) = 150 \left(\frac{1-\epsilon}{D_p G_o / \mu}\right) + 1.75 \quad (9)$$

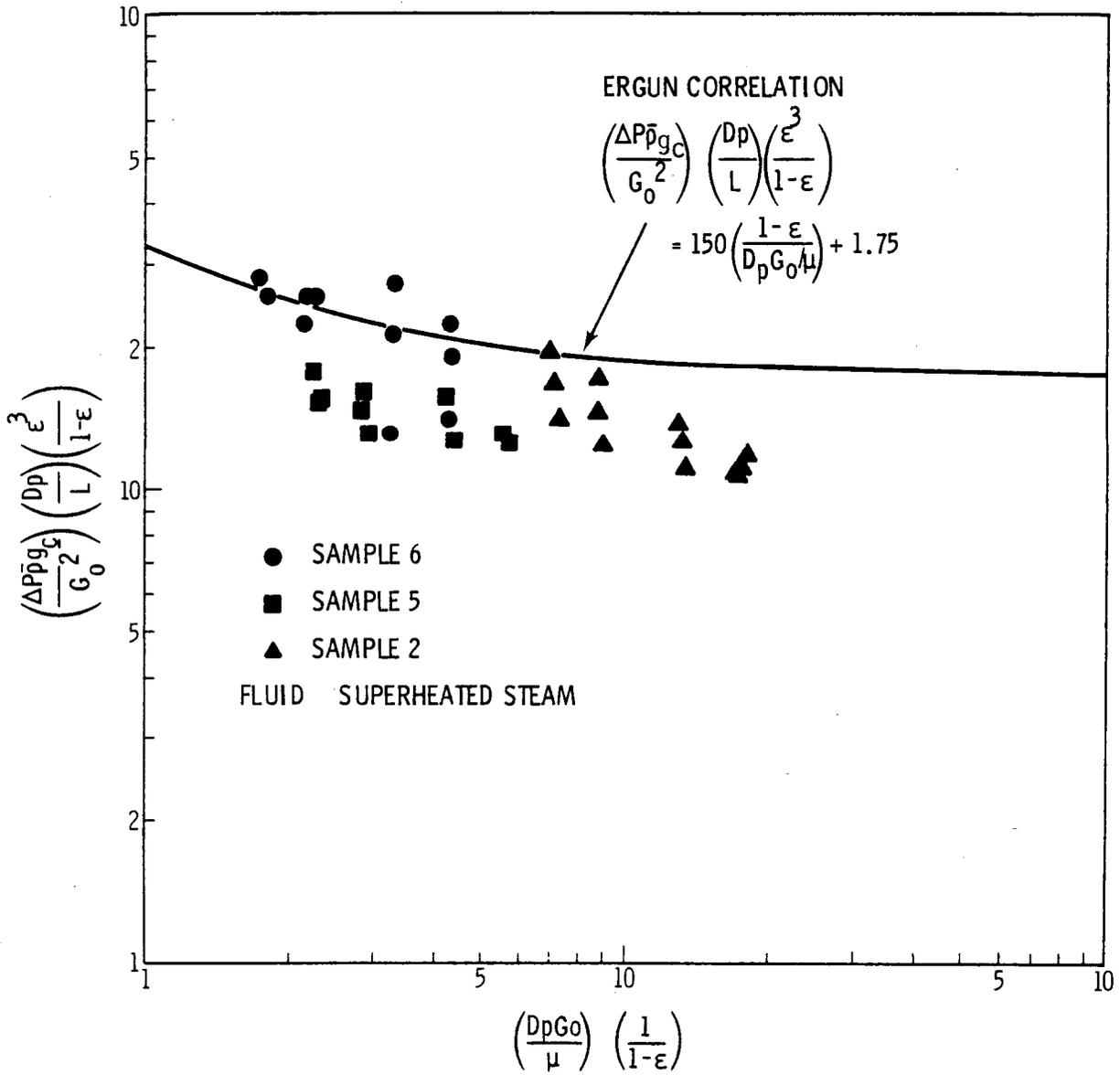


FIGURE 6. Pressure Drop Correlation for Packed Beds - Superheated Steam

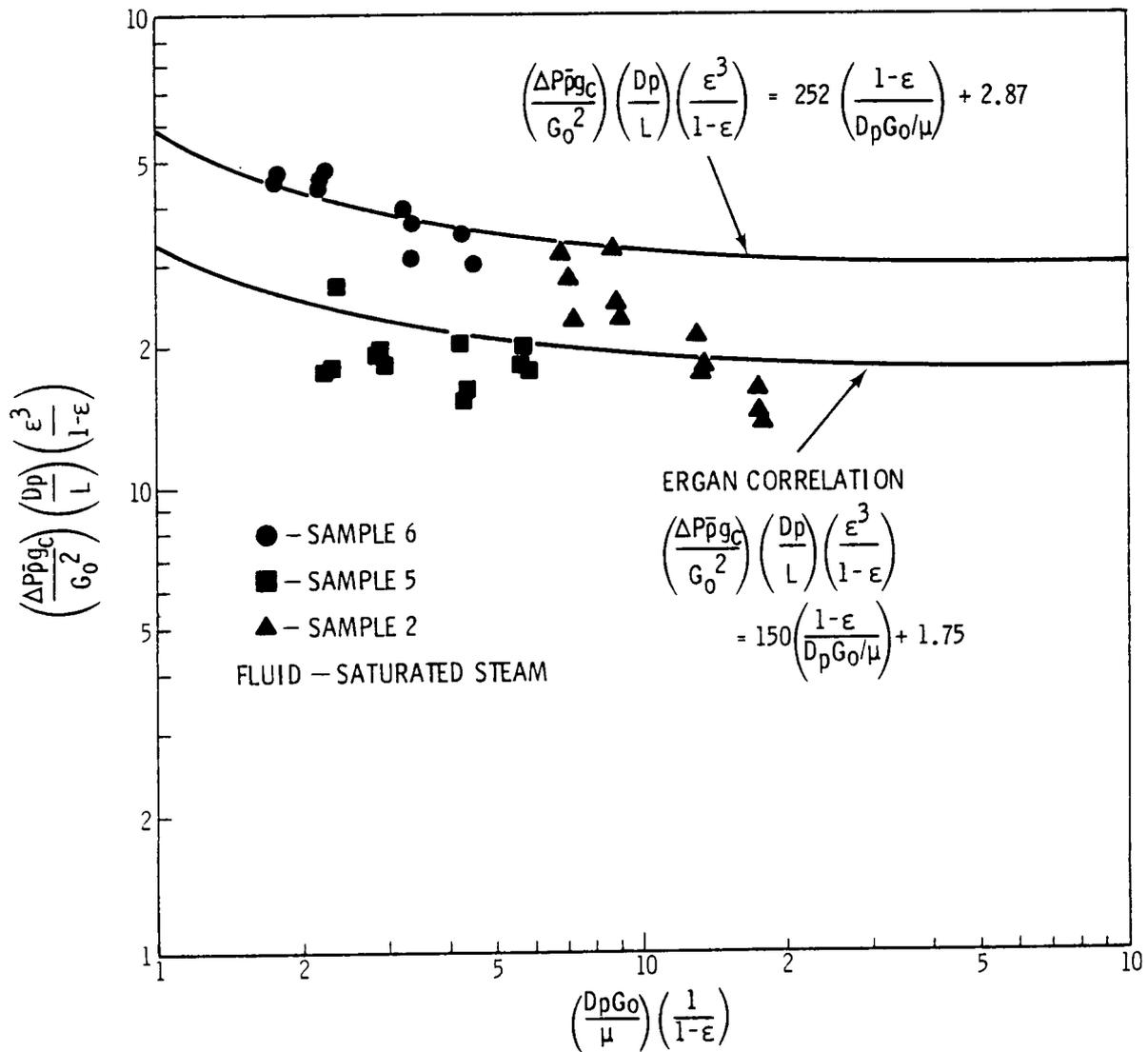


FIGURE 7. Pressure Drop Correlation for Packed Beds - Saturated Steam

where  $\Delta P$  = pressure drop, lb/ft<sup>2</sup>

$\bar{\rho}$  = density mass of fluid, lb/ft<sup>3</sup>

$G_0$  = mass flow rate based on an empty tube, lb/ft<sup>2</sup> sec

$D_p$  = particle size, ft

$L$  = bed length, ft

$\epsilon$  = void fraction

$\mu$  = viscosity, lb/ft sec

$g_c = 32 \text{ ft/sec}^2$

The Ergun correlation was developed from data on the pressure drop of noncondensable gases across packed beds. Data obtained from these series of experiments are presented in Appendix II.

Again, the points plotted on Figure 6 are for clean superheated steam flowing through clean carbon beds. It was concluded that the Ergun equation probably predicts a slightly high pressure drop for the conditions used in this work.

Figure 7 is a similar plot, except data points represent measurements obtained with steam at saturation conditions. An upward shifting of the data points is evident, presumably due to random condensation of steam and partial blockage of the flow path through the bed. Based on the data obtained in this work, a safely conservative prediction of pressure drop for saturated steam pressure drop across carbon beds would be described by:

$$\left(\frac{\Delta P \bar{\rho} g_c}{G_0^2}\right) \left(\frac{D_p}{L}\right) \left(\frac{\epsilon^3}{1-\epsilon}\right) = 252 \frac{1-\epsilon}{D_p G_0 / \mu} + 2.87 \quad (10)$$

It was observed that an increase in bed pressure drop occurred with continuing flow of steam, both during experiments to determine catalyst performance and with only clean steam flowing. This pressure drop increase was generally in the range of seven to eight percent and was always less than 10 percent. This effect was attributed to tighter bed packing with continued differential force as no changes in particle sizes were observed.

## Effects of Noncondensable Gases and Condensate

Geothermal steam contains noncondensable gases such as  $\text{CO}_2$ ,  $\text{H}_2$ ,  $\text{CH}_4$  and  $\text{NH}_3$  other than  $\text{H}_2\text{S}$  as shown in Table 1. Geothermal steam may also contain boric acid and traces of heavy metal vapors such as mercury and arsenic. Although their concentrations in geothermal steam are low, the effect of these gases on the capacity and activity of the catalyst could be critical. For example, it is known that heavy metals may be permanently adsorbed or react with activated carbon. It has been also reported<sup>(2)(6)</sup> that the presence of  $\text{NH}_3$  in a gas stream containing  $\text{H}_2\text{S}$  may sometimes enhance removal of  $\text{H}_2\text{S}$  by the oxidation reaction with activated carbon.

To carry out the investigation the aforesaid noncondensable gases were introduced to the simulated steam one at a time, and finally, a mixture of all gases was added to the system. In these experiments carbon Sample 5, was chosen as the catalyst because this brand gave good  $\text{H}_2\text{S}$  removal results in past experiments. The effect of these gases on the  $\text{H}_2\text{S}$  removal capacity was determined by the measurement of inlet and outlet  $\text{H}_2\text{S}$  concentrations as well as comparison of breakthrough times.

Addition of  $\text{CO}_2$ ,  $\text{H}_2$  and  $\text{CH}_4$  to the steam was accomplished without difficulty; however, introduction of  $\text{NH}_3$  to the system was difficult. Difficulties included an unstable supply of  $\text{NH}_3$ , and plugging of the gas supply line. A cylinder of liquid anhydrous ammonia was used as the source of  $\text{NH}_3$ . The vapor pressure of ammonia at the room temperature is about 125 psia, which is close to the steam pressure. The low pressure of the ammonia supply source was one reason that ammonia flow was not stable. Also, corrosion of the float (ball) in the rotameter by wet ammonia (contaminated with water vapor) and plugging of the ammonia line in the flow-meter occurred from time to time. When the gas mixing station contained moisture, the plugging of the line became worse. It was also found that when ammonia and  $\text{H}_2\text{S}$  were introduced into the system through the same line, the chance of plugging of the gas supply line increased. These problems were eventually solved by heating the cylinder containing ammonia with a hot air gun and by introducing ammonia separately from other noncondensable gases. Ammonia was premixed with air before it was introduced to the system.

In the gas analysis problems were also encountered due to the presence of ammonia or ammonium hydroxide in the sample. It was later learned that the iodometric titration method, as described in Appendix I, is not accurate at high pH. This problem was overcome by adding an extra amount of hydrochloric acid to the iodine solution before it was poured into the steam sample.

Results of the investigation of effects of noncondensable gases and on the H<sub>2</sub>S removal process are summarized in Table 5 and discussed in the following paragraphs:

- a. Carbon dioxide, methane and hydrogen have no noticeable effect on H<sub>2</sub>S removal efficiency and capacity of the catalyst (Sample 5) used in this study even at concentrations twice as high as the typical concentrations of Geysers' steam. In those runs (runs 138, 140 and 142) where gas concentrations were doubled, breakthrough did not occur even after 13 or 13.5 hours of the steam treatment operation. This showed that the performance of the catalytic oxidation process was not affected by the presence of CO<sub>2</sub>, H<sub>2</sub> and CH<sub>4</sub>.
- b. After the problems of introducing ammonia into the reaction system had been solved, runs with the typical NH<sub>3</sub> concentration (Run 151) and with double the NH<sub>3</sub> concentration (Run 145) were made. It was found, from these two runs, that the presence of ammonia in the steam did not have any harmful effect on the performance of the oxidation process. At the end of each run, however, we found that some white powdery material was formed on the surface of the catalyst. Sometimes, because of the formation of this white material, the discharge of spent carbon from the reactor became difficult. A plunger was required to push the spent carbon out of the reactor.

As discussed previously, the formation of (NH<sub>4</sub>)<sub>2</sub>SO<sub>4</sub> as in Reaction (8) was suspected. To identify the chemical constituents of the powdery material formed on the surface of the carbon, the

TABLE 5. Effects of Noncondensable Gases on Catalyst Performance

Run No.	Carbon	Steam Flow Rate-lb/ft <sup>2</sup> -sec	Temperature °C	Duration Hr	Inlet Composition - ppm					Average Outlet H <sub>2</sub> S Content-ppm
					H <sub>2</sub> S	CO <sub>2</sub>	H <sub>2</sub>	CH <sub>4</sub>	NH <sub>3</sub>	
137	No. 5	42.2	190	10	218	6,200	-	-	-	3.5
138	No. 5	41.2	188	10	200	12,400	-	-	-	2.3
139	No. 5	40.5	190	10	210	-	52	-	-	3.7
140	No. 5	39.4	190	13	220	-	104	-	-	3.1
141	No. 5	40.5	190	10	227	-	-	385	-	2.9
142	No. 5	39.4	189	13	200	-	-	790	-	2.0
145	No. 5	40.0	190	9	193	-	-	-	110	6.5
146	No. 5	40.0	185	5.5	156	12,400	104	290	110	5.2
149	No. 5	40.0	185	10	195	12,400	104	790	110	3.4
150	No. 5	40.0	187	10	215	6,200	52	385	55	2.4
151	No. 5	40.0	188	10	215	-	-	-	55	2.5
152	No. 1	40.0	188	10	216	12,400	104	790	160	3.0
163	No. 1	46.4	188	27	193	6,200	52	385	55	1.0 0-5 hr 2.0 5-10 hr 4.3 10-15 hr 7.3 15-20 hr 18.5 20-27 hr

spent carbon was first washed with hot water and the washed water was collected by a filtration. The filtrate was evaporated in a Petri dish to recover any solid. A part of the solid (powder) recovered was redissolved in distilled water. Barium chloride solution was added to the solution to check for the presence of  $(SO_4)^{--}$  ions. Results of the test showed positive. Also, when the solution was heated with concentrated sodium hydroxide solution, gaseous ammonia was released from the solution. This confirmed that the white powder formed on the surface of spent carbon was  $(NH_4)_2SO_4$ . This conclusion also means that whenever  $H_2S$  and  $NH_3$  are present in geothermal steam, the formation of  $(NH_4)_2SO_4$  is possible as a result of the catalytic oxidation reaction. The recovery of  $(NH_4)_2SO_4$  for other usages such as fertilizer during the catalyst regeneration step will depend on the quantity of  $(NH_4)_2SO_4$  formed and the economic situation.

- c. In runs 146, 149, 150, 152 and 163 all the noncondensable gases were mixed with the steam and catalyst performance was examined. In runs 150 and 163 the typical concentrations of noncondensable gases were used, and in runs 149 and 152 twice as high as the typical gas concentrations were used.

Activated carbon used in runs 149 and 150 was Sample 5; runs 152 and 163 employed Sample 1. From these experiments the same conclusion as in the previous cases, where noncondensable gases were introduced to the system one at a time, was drawn. The presence of all noncondensable gases in the simulated steam did not affect the performance of the catalytic oxidation process at all. A 27-hour run using Sample 1 as the catalyst was made in Run 163. The outlet  $H_2S$  concentration at the end of the 27-hour run was 17.62 ppm.

Some indication had been observed that catalyst effectiveness diminished when the steam being treated approached saturation. The presence of a liquid water film which covers the active sites of activated carbon was thought to be the reason. To further investigate this phenomenon, the presence of liquid

water in the reactor packed with activated carbon was attempted by using the electrical properties of the bed. A coaxially embedded stainless rod was used as one electrode, and the wall of the reactor was served as the other electrode. The resistance between these two electrodes was measured. When carbon was saturated or partially saturated with water, the resistance of the bed would be expected to be low; on the other hand, when the bed was dry or the superheated steam was passing through the bed, the resistance of the bed would be high.

Results of catalyst bed resistance measurements for investigating the ineffectiveness of H<sub>2</sub>S removal from saturated steam are summarized in Table 6. It clearly shows that whenever the bed resistance decreases, the outlet H<sub>2</sub>S concentration increases. In other words, when saturated steam passes through the bed, the electrical resistance of the bed decreases, and the outlet H<sub>2</sub>S concentration increases. Possible reasons for the failure to remove H<sub>2</sub>S from saturated steam by the catalytic oxidation process are as follows:

TABLE 6. Carbon Bed Electrical Resistance Measurements

	Temperature °F	Pressure psig	H <sub>2</sub> S Conc.		Bed Resistance ohm
			Inlet ppm	Outlet ppm	
Saturated	324	80	188	85.3	3.5
Superheated	334	80	197	1.2	43.1
Saturated	338	100	237	129.4	2.4
Superheated	348	100	212	0.5	44.6
Saturated	350	120	255	172.1	2.0
Superheated	360	120	220	0.8	43.5

a. A Loss in Active Sites of the Catalyst

When the saturated steam condenses on the surface of the catalyst, it forms a liquid film which covers all or part of the available active sites of the catalyst which need to be occupied by reactants before

they can react with each other.<sup>(13)</sup> As a result of this blinding of active sites, the effectiveness of the catalytic oxidation process decreases drastically.

b. Low Concentrations of Gaseous Reactants in the Liquid Film Covering the Surface of the Catalyst

A calculation of the equilibrium concentration of gaseous reactants in the water film shows that concentrations of  $H_2S$  and  $O_2$  are about ten thousandths of the concentration in the gas phase. Even assuming that the presence of liquid water does not affect catalyst activity, the reduction of the concentrations of reactants will undoubtedly reduce the rate of oxidation. Therefore, the low concentration of reactants at the liquid film-catalyst interface is one of the reasons that the process fails to work in the saturated steam regime.

c. Decrease of Oxygen Solubility in Water Due to the Accumulation of  $H_2SO_4$  in the Liquid Film

As shown in Reaction (6),  $SO_3$  can be produced in the oxidation system as a result of side reactions. When a water film forms at the surface of the catalyst,  $SO_3$  further reacts with water to form  $H_2SO_4$  as shown in the following reaction:



Komiyama and Smith<sup>(14)</sup> found that the solubility of oxygen in water decreased as the concentration of  $H_2SO_4$  in water increased. The presence of  $H_2SO_4$  has been actually detected in our experiments. The color of the activated carbon impregnated with copper compounds changes from black to blue  $CuSO_4 \cdot 5H_2O$  as the  $H_2S$  removal process proceeded. A decrease in the oxygen concentration due to (1) the solubility of oxygen in water and (2) accumulation of  $H_2SO_4$  in the water film formed by the condensation of saturated steam affects the reaction rate of the oxidation reaction.

## Oxygen Requirements

According to Reaction (1) for one mole of  $H_2S$  being converted to sulfur one half mole of oxygen is required stoichiometrically. However, in order to have a more complete removal of  $H_2S$  from geothermal steam, oxygen in excess of the stoichiometric requirement may be introduced. Any oxygen in excess of the stoichiometric requirement will pass through the reactor with the treated steam and enter the turbine power generator. Corrosion problems in the power generating equipment, especially at high temperatures and in the presence of liquid water, may result. The amount of oxygen which maximizes conversion of  $H_2S$  to sulfur and yet not cause significant oxygen corrosion problem must be sought. To determine the optimal oxygen requirement, the bench scale system was operated with different levels of oxygen concentration, and the effectiveness of the  $H_2S$  removal was indicated by the  $H_2S$  concentration in the reactor outlet stream. The unreacted oxygen was determined either directly by a gas chromatographic analysis or indirectly calculated from the  $H_2S$  conversion.

In earlier work when limitations of operating variables were investigated, it was found that oxygen concentrations somewhere between 1.3 and 1.6 times the stoichiometric requirement were needed for effective removal of  $H_2S$ . With a more careful control of the operating variables for the determination of oxygen requirement two runs using carbon, Sample 1 and Sample 5 were carried out. Data from these runs are summarized in Tables 7 and 8.

In Run 157, when the air flow rate decreased from 57 to 30 ml/min (1.48 to 1.02 times of the stoichiometric requirement) the outlet concentration of  $H_2S$  increased from less than 1 ppm to 19.81 ppm. There was a sharp jump in the outlet concentration of  $H_2S$  when the oxygen supply was changed from 1.14 to 1.02. In Run 158, the change in the oxygen supply from 1.85 to 0.92 caused the outlet  $H_2S$  concentration to change from 0.83 to 27.65 ppm. However, the change in the outlet concentration of  $H_2S$  with respect to the change in oxygen supply is smoother in Run 158 than in Run 157.

TABLE 7. Minimum Oxygen Requirement for Carbon Sample 5

Run 157

Air Flow Rate	57	50	45	40	35	30
Inlet H <sub>2</sub> S ppm	229	188	190	197.3	182	188.1
H <sub>2</sub> S Flow Rate ml/min	16.19	13.29	13.43	13.95	12.87	13.30
H <sub>2</sub> S/O <sub>2</sub>	1.37	1.27	1.42	1.66	1.75	2.11
Theoretical Oxygen Flow Rate, ml/min	8.10	6.65	6.77	6.98	6.44	6.15
O <sub>2</sub> /Theo. O <sub>2</sub>	1.48	1.58	1.40	1.20	1.14	1.02
Outlet H <sub>2</sub> S, ppm	1	0.85	0.83	0.42	4.6	19.87
Outlet O <sub>2</sub> in Outlet Noncondensable Gas, %	3.2	2.5	2.0	1.6	1.1	0.75
Condensate Dissolved Oxygen, ppm	2.2	2.1	2.4	1.7	0.4	0.4

TABLE 8. Minimum Oxygen Requirement for Carbon Sample 1

Run 158

Air Flow, ml/min	57	50	45	40	35	30
Inlet H <sub>2</sub> S, ppm	183	198	192.2	194.9	191.6	193
H <sub>2</sub> S Flow Rate ml/min	12.94	14.00	13.59	13.78	13.55	13.65
Theoretical O <sub>2</sub> Req.	6.47	7.00	6.80	6.89	6.78	6.33
O <sub>2</sub> /Theo. O <sub>2</sub>	1.85	1.50	1.39	1.22	1.08	0.92
Outlet H <sub>2</sub> S, ppm	0.83	4.14	6.79	13.20	20.92	27.65
Outlet O <sub>2</sub> in Noncon- densible Gas, %	4	2.5	1.75	1.5	7.1	1.0
Dissolved Oxygen in Condensate, ppm	2.7	2.3	1.1	0.7	0.5	1.8

The results of these experiments indicate that the minimum amount of oxygen required for an effective  $H_2S$  removal from geothermal steam depends on the kind of carbon used in the process. Data showed that the minimum oxygen requirement for the catalytic oxidation reaction using Sample 1 is about 1.5 times the stoichiometric requirement and for Sample 5 about 1.14.

Although outlet oxygen concentrations were determined by a gas chromatograph and concentrations of dissolved oxygen by an electrode probe, these numbers are not very reliable. In order to determine how much unreacted oxygen may be present in the reactor effluent, calculations were made for the typical geothermal steam shown in Table 9. A total conversion of  $H_2S$  to sulfur was assumed in these calculations. The result of the calculations showed that the outlet oxygen concentration is zero ppm at the stoichiometric requirement and 98 ppm at two times the stoichiometric requirement. In the actual operation of the oxidation process, the outlet oxygen concentrations are somewhere between 40 to 70 ppm.

### Catalyst Regeneration

Sulfur is deposited on the surface of activated carbon as a result of the oxidation reaction. Consequently, the activity of the catalyst decreases gradually with continued exposure to steam containing  $H_2S$ . When the concentration of  $H_2S$  in the treated steam becomes too high, the spent carbon must be either discarded or regenerated. Regeneration of the catalyst involves the recovery of sulfur and reactivation of the catalyst. Though the activated carbon catalyst used in the proposed process is relatively inexpensive, regeneration and reuse of it would undoubtedly reduce the operating cost and waste disposal problems.

Various sulfur recovery techniques are available. Vacuum distillation, inert gas (or steam) entrainment, solvent extraction and conversion of sulfur to volatile compounds are a few possibilities.

In vacuum distillation sulfur is recovered by evaporating the deposited sulfur under a subatmospheric pressure. The normal boiling point of sulfur is  $445^{\circ}C$ . This means that sulfur evaporates at a temperature lower than  $445^{\circ}C$  under a vacuum condition. The sulfur vapor can be condensed in a

TABLE 9. Estimated Mass Balance and Oxygen Concentration in Treated Steam

	Raw Steam Moles	Oxygen Supply (Times Stoichiometric Requirement)					
		2.0	1.5	1.4	1.3	1.2	1.0
		Treated Steam, Moles					
CO <sub>2</sub>	141.1	141.1	141.1	141.1	141.1	141.1	141.1
H <sub>2</sub>	26.0	26.0	26.0	26.0	26.0	26.0	26.0
H <sub>2</sub> S	6.1	-	-	-	-	-	-
CH <sub>4</sub>	24.1	24.1	24.1	24.1	24.1	24.1	24.1
NH <sub>3</sub>	3.2	3.2	3.2	3.2	3.2	3.2	3.2
N <sub>2</sub>	3.3	26.3	20.5	19.4	18.2	17.1	18.4
H <sub>2</sub> O	55,166.6	55,172.7	55,172.7	55,172.7	55,172.7	55,172.7	55,172.7
O <sub>2</sub>	0.0	3.1	1.5	1.2	0.9	0.6	0.0
Oxygen Concentration, ppm	0.0	99.0	49.0	39.0	29.0	20.0	0.0

condenser operated above the melting point of sulfur. The sulfur so recovered is in the form of a liquid.

Inert gas (or steam) entrainment may be accomplished by passing hot inert gas or superheated steam (whose temperature is higher than the boiling point of sulfur) through the bed of spent catalyst. Due to the flow of the inert gas or steam, the vapor of sulfur is entrained from the bed and recovered in a condenser as in the vacuum distillation method.

Solvent extraction is a simple unit operation used in many chemical industries. In the solid-liquid extraction a solid solute (sulfur in this case) is leached from an insoluble residue (activated carbon) into a liquid solvent phase. The solution is subsequently separated from the solid, and the solute is recovered by precipitation. The resulting mother liquid can be reused for the next batch extraction either with or without further purification.

Sulfur solubilities in selected solvents are shown in Figure 8.<sup>(15)</sup> The solubility of sulfur in carbon disulfide is 52 gm/100 gm of carbon disulfide at its normal boiling point (46°C). Anhydrous liquid ammonia has an unusual solubility curve; sulfur solubility decreases as temperature increases. To use ammonia as a solvent for sulfur extraction, a low temperature and/or pressurized system would be required. Other solvents such as quinoline, ethylene dibromide,  $\beta$ -naphthanol, toluene and ethylene chloride, whose sulfur solubility-temperature curves have steep slopes, could be used for the extraction of sulfur. Sulfur extraction can be accomplished at high temperature and the recovery of sulfur can be done at temperatures lower than that of the extraction. Many solvents are toxic and/or hazardous to the environment. Therefore, in selecting a solvent for sulfur extraction, handling hazards must be considered as well as carbon regeneration effectiveness.

After sulfur extraction, the spent carbon may not be directly reusable because some solvent may be retained in its pores. The amount of adsorbed solvent (residual solvent) depends upon the kind of solvent used for the extraction and the kind of carbon being treated. Before the carbon can be reused, the residual solvent must be removed from the carbon. It may also be

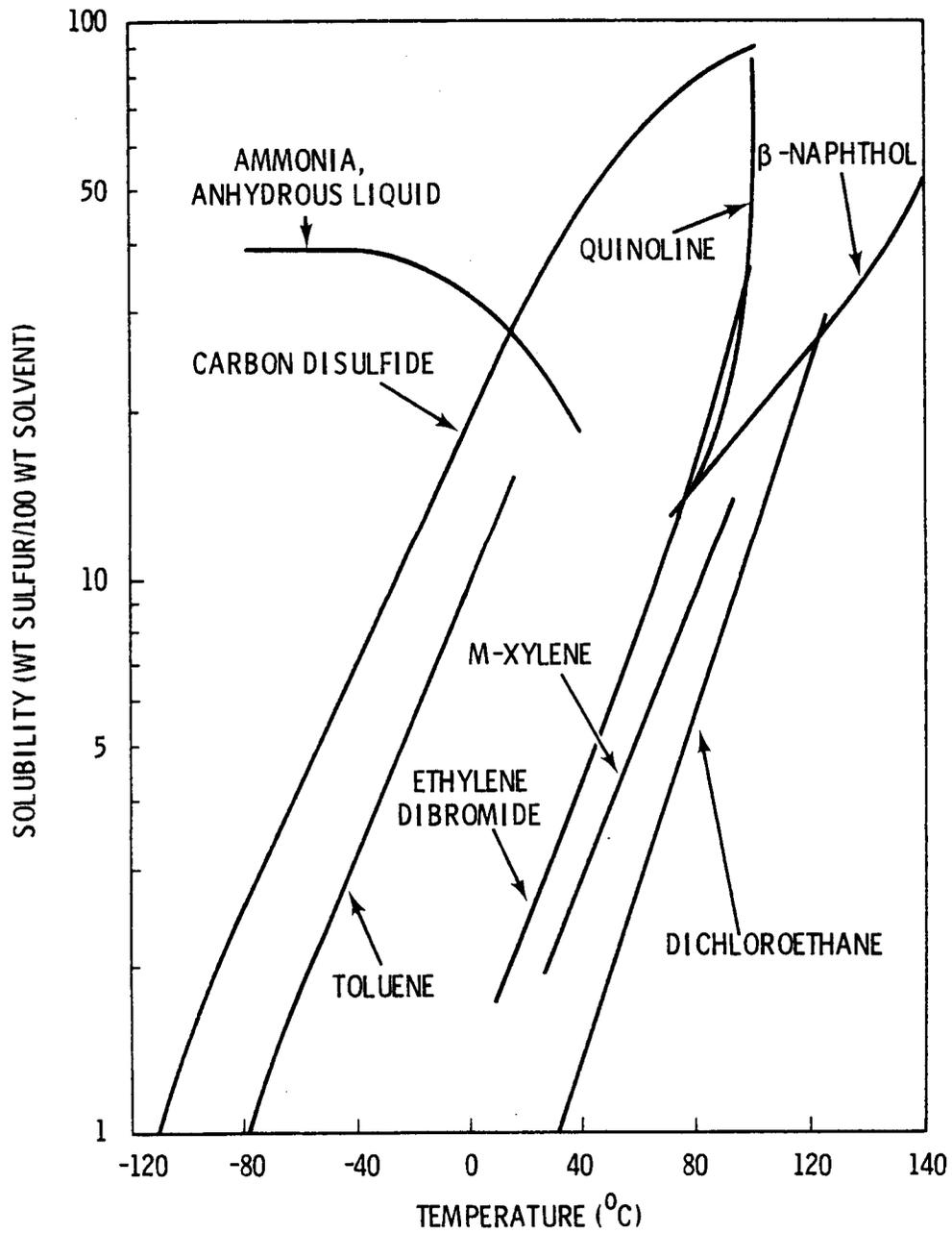


FIGURE 8. Solubility of Rhombic Sulfur in Solvents

necessary to reactivate the carbon to restore its activity. The reactivation of the carbon can be accomplished by treating the sulfur extracted carbon with an activation agent such as steam, oxygen, carbon monoxide, carbon dioxide or a mixture of these gases.

The activity of regenerated carbon, the amount of sulfur recovered, the amount of energy required and the cost of regeneration have been studied. To study the activity of successively regenerated carbon, carbon was repeatedly regenerated and reused, and any change in the breakthrough time of the successive runs was compared and evaluated.

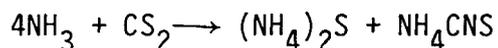
Two regeneration methods for spent carbon have been studied: solvent extraction and steam distillation. Solvent extraction has been emphasized because it is the least energy intensive. Based on the solubility data presented in Figure 8, we chose  $\text{CS}_2$ ,  $(\text{NH}_4)_2\text{S}$ , tetrachloroethylene (TCE), and dichloroethane (DCE) for study. The results of catalyst regeneration experiments will be discussed in the following paragraphs in chronological order.

In Run R1-1, spent carbon (Sample 5) from several previous runs was used. Sulfur recovery was accomplished by extraction using  $\text{CS}_2$  as a solvent. After sulfur had been extracted, the carbon was heated in a vacuum oven for three hours at  $100^\circ\text{C}$ . Before the carbon was reused for  $\text{H}_2\text{S}$  oxidation, it was purged with superheated steam ( $210$  to  $220^\circ\text{C}$  at  $100$  psig) for about 2-1/4 hours to further remove the residual solvent. The subsequent  $\text{H}_2\text{S}$  removal run ( $40$  lb/ft<sup>2</sup> min at  $200$  ppm  $\text{H}_2\text{S}$ ) ran for  $14.8$  hours and the  $\text{H}_2\text{S}$  concentration in the outlet stream at the end of the run was  $17.4$  ppm. This run using the regenerated carbon was very encouraging.

At the completion of this run (R1-1), sulfur removal by steam distillation was attempted. The spent carbon in the reactor was purged with superheated steam ( $250$  to  $280^\circ\text{C}$ ,  $100$  psig). About  $15$  kg ( $33$  lb) of superheated steam was used. During the steam purging, sulfur was entrained from the bed and collected in the gas-liquid separator as a colloidal solution. The sulfur was recovered from the solution by treating the suspension with a small amount of alum and then by filtration. A total of

4.98 g of sulfur was recovered. The carbon so regenerated was used again in Run R1-2. Run R1-2 lasted only for about 3.5 hours. The reason for the short run could be attributed to the incomplete removal of sulfur from the spent carbon by the steam distillation method. The Run R1-2 spent carbon was then extracted with CS<sub>2</sub>. As a result of the extraction, about 6.18 g of sulfur was recovered. A part of this sulfur was left over from the steam distillation. The sulfur recovery using the steam distillation method was therefore judged to be ineffective.

A second series of catalyst regeneration experiments was conducted with the Sample 1 spent carbon. The spent carbon was first subjected to a CS<sub>2</sub> extraction. Instead of using steam to purge out the adsorbed solvent, gaseous NH<sub>3</sub> was introduced from top to bottom. The gas effluent was bubbled through a flask containing distilled water. Because of the exothermic reaction of NH<sub>3</sub> with the residual solvent, CS<sub>2</sub>, the temperature of the carbon bed increased gradually. As the reaction zone moved from the top to the bottom of the column, so did the temperature profile of the bed. At the end of the reaction, the color of the water in the flask changed to orange. The following reaction was thought to occur in the bed.



A part of the gaseous (NH<sub>4</sub>)<sub>2</sub>S was believed to be absorbed in the water. Both (NH<sub>4</sub>)<sub>2</sub>S and NH<sub>4</sub>CNS are water soluble; therefore, after the reaction was over, carbon in the bed was washed with water to remove these products.

The NH<sub>3</sub> treatment method was used to regenerate the carbons used in Runs 113 and 163. The regenerated carbons were used in Run 113-1 and 163-1, respectively. Run 113-1 lasted for 12.3 hours and Run 163-1 lasted for 15 hours. In this regeneration scheme, the regular steam treatment prior to the H<sub>2</sub>S removal experiment was eliminated.

A third type of regeneration was carried out by extraction of sulfur with an aqueous solution of (NH<sub>4</sub>)<sub>2</sub>S. Carbon used in Run 113-1 was slurried into aqueous (NH<sub>4</sub>)<sub>2</sub>S solution in an Erlenmeyer flask. The flask was

stirred in a shaker for about 30 minutes. After the shaker stopped and carbon particles were settled from the solution, the S -  $(\text{NH}_4)_2\text{S}$  solution was decanted from the flask and another batch of fresh aqueous solution of  $(\text{NH}_4)_2\text{S}$  was added to the carbon in the flask. The same extraction procedure was repeated eight times until 1000 ml of the aqueous solution ( $(\text{NH}_4)_2\text{S} / \text{H}_2\text{O} = 6/4$ ) was used. This regenerated carbon was used in Run 113-2. The  $\text{H}_2\text{S}$  removal experiment using this regenerated carbon lasted only for about 6.33 hours before breakthrough. A possible reason for the short run was that the sulfur extraction using  $(\text{NH}_4)_2\text{S}$  solution was ineffective. A longer, more thorough extraction using more solvent might have improved performance.

In Run 138-1, sulfur extraction by TCE was attempted. The spent carbon, Sample 5, used in Run 138 was subjected to TCE extraction in a Soxhlet apparatus. The extraction recovered 12.65 g of sulfur. Before the carbon was reused for  $\text{H}_2\text{S}$ -steam treatment, it was treated with clean steam (about 4 kg). The subsequent  $\text{H}_2\text{S}$  removal experiment was run for a total of 13 hours with an interruption between the working shifts. At the end of 13 hours the outlet  $\text{H}_2\text{S}$  concentration was about 12.96 ppm. After Run 138-1, the carbon was regenerated by TCE using the same technique as before. However, during the steam treatment, a strange smelling gas was noticed. It was subsequently learned that TCE could be converted to phosgene type of gas under the experimental conditions (and in the presence of oxygen and carbon). A private communication with Dow Chemical Solvent Division confirmed the possibility of phosgene formation. Because of the possible formation of this toxic gas, the use of TCE for sulfur extraction was not further pursued.

In the next series of experiments, DCE was used as a solvent for extraction. Carbon from Run 150 was extracted with DCE. After the extraction, the carbon was dried in an oven at  $100^\circ\text{C}$  overnight. The dried carbon (80 g) was charged into the reactor and purged with steam. A normal  $\text{H}_2\text{S}$  removal experiment, Run 159-1, was started when the steam treatment had been finished. The run lasted for 13.75 hours before breakthrough. After Run 159-1, the spent carbon was removed from the reactor and packed in a glass column wound with a heating tape. The extraction was accomplished by a

batchwise technique. About 120 ml of the solvent solution (DCE saturated with sulfur at 40°C) was charged into the carbon-packed glass column. The column temperature was kept at 80°C by controlling a Variac connected to the heating tape. The carbon was soaked with the warm solution for about 30 minutes. The solution was then drained from the column and collected in a flask. This was repeated until a total of 600 ml of the saturated solution was used. The carbon was then dried in an oven. In Run 159-1a, this DCE extracted carbon (Sample 5) was used. This run lasted less than two hours before breakthrough. The ineffectiveness of sulfur removal by the batchwise solvent washing method was thought to be the reason for the short run. After Run 159-1a DCE was again used as the solvent and sulfur was extracted from the spent carbon in a Soxhlet. The sulfur-free carbon was dried and was used in Run 159-2. By following the regular procedure, Run 159-2 continued for 12.5 hours before breakthrough. This showed that the incomplete sulfur removal discussed in the above was the main reason for the failure of Run 159-1a.

The spent carbon of Run 159-2 was extracted with DCE in Soxhlet again and was used in Run 159-3. The data for Run 159 series are shown in Figure 9.

The above results are summarized as follows:

- $\text{CS}_2$  is a good solvent for the removal of sulfur from spent carbon. However, carbon adsorbs  $\text{CS}_2$  which must be removed before it can be reused. Steam or ammonia treatment can be used for this purpose. However, steam treatment requires a large amount of steam (230 lb steam/lb of spent carbon). Ammonia treatment would involve washing the spent carbon with water to remove water soluble byproducts such as ammonium sulfate,  $\text{CS}_2$  extraction to recover sulfur and treatment of the  $\text{CS}_2$  contaminated carbon with ammonia to form  $(\text{NH}_4)_2\text{S}$  and  $\text{NH}_4\text{CNS}$ . After washing with water, the carbon can be reused. Sulfur, recovered from the solvent by evaporation of  $\text{CS}_2$  would be the main product. Ammonium sulfate could also be produced if sufficient  $\text{NH}_3$  is present in the raw steam. Ammonium sulfide and ammonium thiocyanate would contaminate the water used for washing following the ammonia treatment. The use of  $\text{CS}_2$  as a regeneration solvent is also undesirable because it is toxic and special handling procedures would have to be devised to avoid hazards to operating personnel and emissions to the atmosphere.

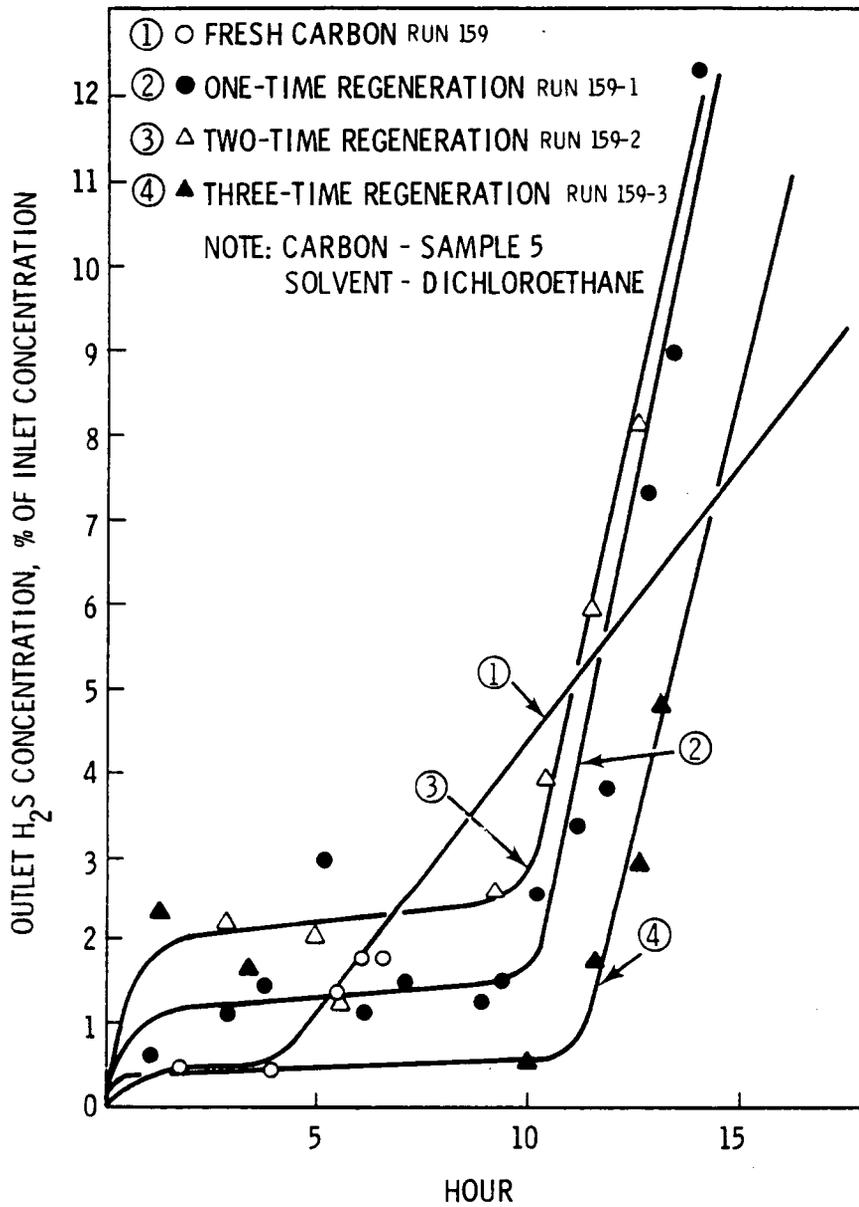


FIGURE 9. Outlet H<sub>2</sub>S Concentration versus Time for Fresh and Regenerated Carbon Sample 5

- Steam distillation entrains some sulfur from spent carbon. However, to completely remove sulfur from the carbon the quantity of the superheated steam will be too large to be practical.
- Extraction of sulfur from spent carbon with  $(\text{NH}_4)_2\text{S}$ , though described in the literature as a useful method for dissolving sulfur, was not successful.
- The use of tetrachloroethylene (TCE) is impractical as an extracting media because of the formation of phosgene or phosgene derivations.
- Dichloroethane appears to be an excellent candidate as a carbon regeneration solvent. Carbon regenerated by DCE extraction three times gave very good performance. It is apparent that a regeneration method using this solvent and sulfur recovery by solvent evaporation is technically feasible. However, a method involving dissolution of sulfur at an elevated temperature ( $80^\circ\text{C}$ ) with solvent saturated at  $40^\circ\text{C}$  (simulating a solvent from which sulfur had been removed by cooling and precipitation) was not effective. It is possible that this method could still prove practical with more optimum contact times and temperatures.

#### Corrosion Considerations

It has been shown that a slight excess of oxygen, in comparison to stoichiometric requirements, is required to sustain efficient  $\text{H}_2\text{S}$  oxidation in the proposed process. Unreacted oxygen in treated steam may corrode the power generating equipment. Corrosion problems may also be intensified if the steam to be treated contains chlorides. There is also likely to be some sulfate formation which, if washed from the catalyst by steam condensate, may also be detrimental. The available resources for this project prevented experimental study of these effects. However, the general problem was qualitatively considered in the following paragraphs.

Data on the corrosion of steam turbines due to the presence of oxygen in the steam is scarce. However, the corrosion of boiler tubes due to dissolved oxygen in the boiler feed water has been extensively studied.<sup>(16)</sup> Speller reports that "dissolved oxygen is usually the controlling factor in corrosion of iron in water," also, "for low-pressure steam boilers operating under 250

psi without an economizer, the desirable maximum (oxygen concentration) should not exceed about 0.03 ppm (0.02 ml/liter)." From these statements it would appear that corrosion due to the presence of liquid water with dissolved oxygen is more severe than dry steam with oxygen.

In earlier work, it was found that up to 50 ppm of oxygen may be present in steam treated by the proposed process. The maximum amount of dissolved oxygen (the equilibrium concentration) in steam condensate based on 50 ppm unreacted oxygen in the treated steam is estimated as 0.006 ppm. This concentration is about one order of magnitude smaller than the recommended maximum. It may be concluded that accelerated corrosion in carbon steel piping and components, with use of the H<sub>2</sub>S oxidation process, is unlikely. However, this conclusion must be confirmed. The most practical approach would be to perform such a corrosion evaluation in conjunction with the next phase of development--demonstration of the process under realistic field conditions.

### CONCEPTUAL PROCESS DESIGN

The conceptual process outlined in the following paragraphs is based on the use of fixed beds of activated carbon, injection of air to provide the required oxygen, and carbon regeneration by the use of dichloroethane (DCE) solvent. Material and energy balances are derived for a 50 MWe sized plant, based on steam properties typical of the Geysers.

It may be necessary to provide for avoiding condensate in the catalyst beds if the supply steam is at or near saturation. This may be accomplished by adiabatic steam expansion or by adding energy to superheat the raw steam. A detailed comparison of these alternatives is given in Appendix III. However, it is likely that at most locations where the steam source is at saturation there will have been sufficient nonfrictional pressure drop through valves, fittings and filters to provide adequate superheat margin.

### Overall H<sub>2</sub>S Removal Process Flow Diagram

A conceptual process diagram is shown in Figure 10. Basically the process utilizes multiple reactors in a series scheme. A three-reactor system will be used here for explanatory purposes. In Figure 10-a, Reactor 1 and 2

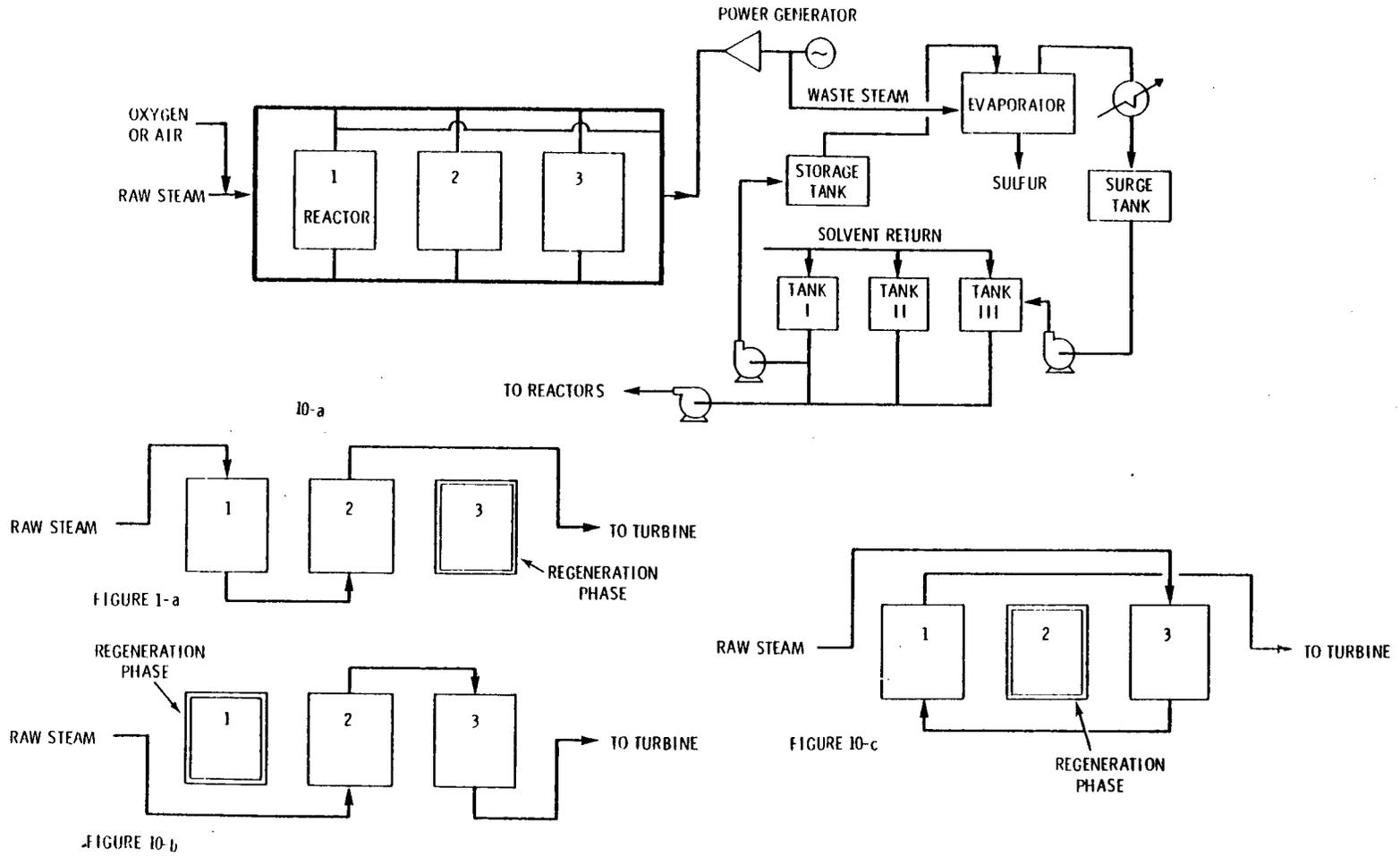


FIGURE 10. Conceptual Process Design

are "on-line" while Reactor 3 is being regenerated. Geothermal steam passes through reactors 1 and 2 to remove  $H_2S$ . In Figure 10-b Reactor 2 and 3 are "on-line" while Reactor 1 is being regenerated. Figure 10-c shows that Reactor 3 and 1 are "on line" and Reactor 2 is being regenerated. With this operating sequence the capacity of the catalyst can be fully utilized while maintaining a high  $H_2S$  removal efficiency. When the length of the catalyst bed used for the  $H_2S$  removal process is as short as in our laboratory experiments (nine to fifteen inches) instead of the reactor in series scheme, a three-parallel reactor system will be used. Two of them will be for  $H_2S$  removal, and one will be for regeneration.

The solvent used for catalyst regeneration is stored in three (or more) tanks. The spent catalyst is first contacted with solvent having the highest sulfur content stored in Tank I. The effluent is pumped to an evaporator where elemental sulfur is recovered. The energy required for solvent evaporation is supplied either from the exhausted steam of a turbine power generator or from treated geothermal steam. The spent catalyst is again extracted with solvents having lower sulfur concentration stored in Tank II. The effluent of the extraction is stored in Tank I. The final step is to extract sulfur by a pure solvent stored in Tank III with the effluent of the extraction stored in Tank II. The pure solvent vapor from the evaporator is recovered in a condenser. The solvent condensate goes to a surge tank and finally is pumped to Tank III.

When a solvent with a sulfur solubility sensitive to temperature variations is used, a high-temperature extraction and a low-temperature sulfur separation will be used. Dichloroethane is a solvent of this type. In the extraction, a hot solvent is pumped to the reactor, and the spent carbon is soaked in the hot solvent for a period of time. The hot solution is then discharged to a crystallizer where the solution is cooled and sulfur crystals form. The sulfur is then separated from the solution by a centrifuge or filter. The cold mother liquid is pumped through a heater and then to the reactor for another run of extraction. The same operation is repeated until most of the sulfur is removed from the spent carbon.

## Material Balance

A material balance for a 50 MWe geothermal power plant using steam having a composition the same as the Geysers (see Table 1) is presented in Figure 11. In order to prepare the material balance, the following assumptions are also made:

- (1) Plant size: 50 MWe
- (2) Steam flow rate: 1,000,000 lb/hr
- (3) Basic data: See Run 163 in Appendix II
  - (a) Overall average outlet  $H_2S$  concentration is 6 ppm.
  - (b) 10% of  $NH_3$  in steam is converted to  $(NH_4)_2SO_4$ .
  - (c) Breakthrough time is 12 hours (although Run 163 lasted for 27.5 hours).

As shown in the figure, the process uses 7.6 tons/day of air (or 1.5 tons/day of oxygen), produces 2.2 tons/day of elemental sulfur and 0.26 tons/day of  $(NH_4)_2SO_4$ . The amount of carbon required, the number of reactors and reactor size was extrapolated from the bench scale data of Run 163. By using the same steam flow rate, space velocity (or residence time) the following calculations were made:

### Amount of Carbon Required and Size of Reactor

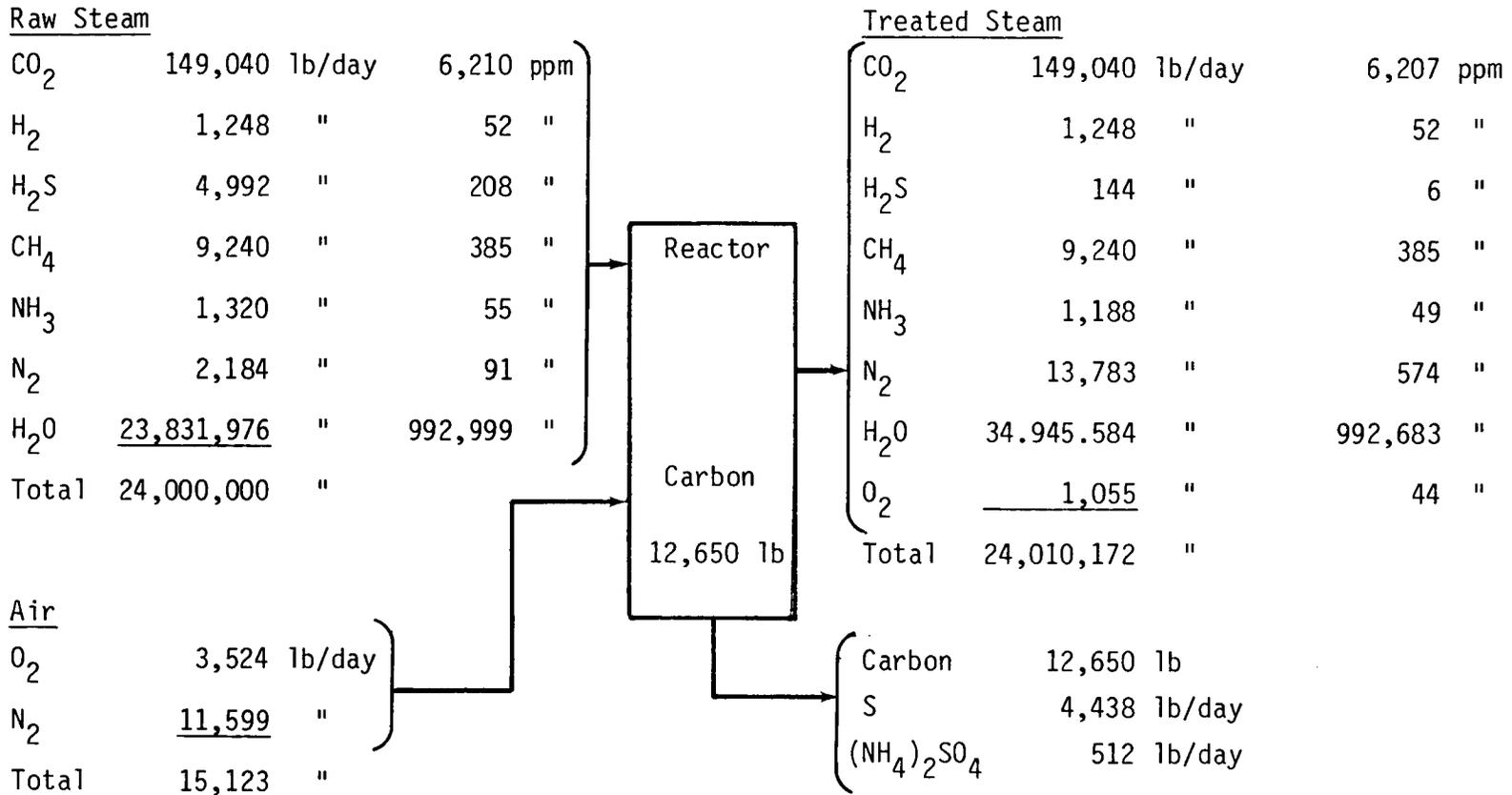
The amount of carbon required is:

$$\begin{aligned} W &= 80g \times \frac{1}{453.6} \frac{lb}{g} \times 1 \times 10^6 \times \frac{1b}{hr} \times \frac{1}{60} \frac{hr}{min} \times \frac{1}{105.4} \frac{min}{g} \times 453.6 \frac{g}{lb} \\ &= 12,650 \text{ lb} = 6.33 \text{ ton} \end{aligned}$$

The total cross section reactor area can be calculated as:

$$\begin{aligned} A &= 1 \times 10^6 \frac{lb}{hr} \cdot \frac{1}{60} \frac{hr}{min} \cdot \frac{1}{105.4} \frac{min}{g} \cdot 453.6 \frac{g}{lb} \times 0.00499 \text{ ft}^2 \\ &= 357.92 \text{ ft}^2 \end{aligned}$$

FIGURE 11. Material Balance of the H<sub>2</sub>S Removal Process for a 50 MWe Geothermal Power Plant



Since the total area is equal to the number of reactors times the cross sectional area of each reactor, the number of reactors and the diameter of the reactor can be calculated as follows:

$$D = \sqrt{\frac{4A}{\pi N}}$$

where D = diameter of reactor

N = number of reactors

The following table shows the relation between the number of reactors and reactor diameter.

<u>Number of Reactors</u>	<u>Diameter of Reactor, ft</u>
1	21.35
2	15.09
4	10.67
6	8.72
8	7.55
10	6.75
12	6.16
14	5.71

If the steam required for a power plant is supplied by seven wells (typical Geysers well output) and each well is connected to three parallel reactors, two for H<sub>2</sub>S removal and one for regeneration, a total of 21 5.7 ft diameter reactors are required.

According to data obtained from Run 163, the depth of the carbon bed is 14.2 inches. If both the inlet and outlet ends of the bed are equipped with six inches of free space for even steam distribution, the overall length of a reactor will be 2.2 ft.

Volume of carbon in each bed is:

$$V_c = (5.71)^2 \times \frac{3.14}{4} \times 2.2 = 56.34 \text{ ft}^3 = 421.4 \text{ gal.}$$

and each reactor contains:

$$W_c = 12650.22/14 = 903.6 \text{ lb of carbon}$$

As a summary, the conceptual H<sub>2</sub>S removal plant will have reactors of following dimensions:

Diameter: 5.7 ft

Length: 2.2 ft

Volume of carbon = 56.34 ft<sup>3</sup>/reactor

Weight of Carbon = 903.6 lb/reactor

Total number of reactors = 21 (14 for H<sub>2</sub>S removal and 7 for regeneration)

#### Amount of Solvent Required

Regeneration will be by pumping a solvent into the reactor and allowing the carbon to soak in the solvent for a period of time. Sulfur is leached from the spent carbon and dissolved in the solvent. The sulfur solution is then drained from the reactor and sent for sulfur recovery. The amount of solvent required for each batch regeneration is calculated as follows:

Assumption:

- (1) Voidage of bed: 0.39
- (2) Porosity volume of carbon = 0.7 cc/g
- (3) The amount of solvent to be pumped into the reactor will be enough to cover the upper surface of carbon bed and fill the free space at the bottom part of the reactor.

The amount of solvent required for each reactor is:

$$V_s = \frac{(5.7)^2 \pi}{4} \left[ \frac{14.2}{1.2} \times 0.39 + \frac{6}{12} \right] + 0.7 \times \frac{453.6}{(30.48)^2} \times 903.6$$

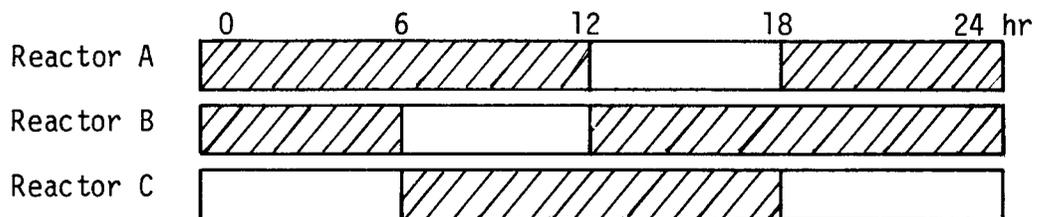
$$= 34.67 \text{ ft}^3 = 260 \text{ gallons}$$

If three consecutive batchwise washings are required, the total volume of solvent will be:

$$V = 260 \times 14 \times 3 = 10,920 \text{ gallons}$$

### Schedule for Carbon Regeneration

The schedule for carbon regeneration for a three-reactor steam treatment unit is shown in the following diagram.



The shaded areas indicate that the reactor is on line for H<sub>2</sub>S removal.

When time equals zero, both Reactor A and B are on line for H<sub>2</sub>S removal; Reactor C starts into regeneration. At six hours Reactor A continues H<sub>2</sub>S removal; Reactor B starts into regeneration; Reactor C finishes regeneration and starts H<sub>2</sub>S removal. At 12 hours Reactor A starts into regeneration; Reactor B finishes regeneration and is ready for H<sub>2</sub>S removal; Reactor C continues H<sub>2</sub>S removal. At 18 hours Reactor A finishes regeneration and is ready for H<sub>2</sub>S removal; Reactor B continues H<sub>2</sub>S removal; Reactor C starts into regeneration. At 24 hours, Reactor A continues H<sub>2</sub>S removal; Reactor B starts regeneration; Reactor C finishes the regeneration and is ready for return to H<sub>2</sub>S removal. At any time two reactors will be on line for H<sub>2</sub>S removal and one reactor will be in the regeneration cycle. Each reactor will be operated 12 hours for the removal of H<sub>2</sub>S and six hours for the regeneration.

### PROCESS ECONOMICS

Preliminary estimates of operating and capital costs were developed for the conceptual process described in the previous section. All essential process features were included. It was assumed that superheating of the steam

would be required to avoid condensation in the catalyst beds. It should be recognized that these cost estimates were derived at a stage of process development where system and equipment details are yet to be defined.

The basis utilized for capital and operating costs were:

- (1) Capital cost items were those for carbon, solvent, major equipment, instrumentation, piping and valves, and offsite facilities.
- (2) The major equipment items are reactors, solvent storage tanks, air compressor, heat exchanger, crystallizer, centrifuge, and superheater.
- (3) 21 oxidizer reactors with dimensions previously given are needed.
- (4) 14 solvent storage tanks with capacity of 400 gallons each are required.
- (5) An air compressor to supply 15,125 lb/day of air to the steam treatment system is needed.
- (6) Two heat exchangers, one for solvent preheating and one for solvent recovery are required.
- (7) The instrumentation cost is five percent of the major equipment cost.
- (8) The cost of piping and valve is ten percent of the major equipment and instrumentation costs.
- (9) The cost of offsite facilities is 50 percent of the sum of major equipment, instrumentation and piping and valve costs.
- (10) The annual depreciation is five percent of total capital investment.
- (11) Four operators, one maintenance person, and one supervisor are required to operate the H<sub>2</sub>S removal plant.
- (12) Energy required for steam superheating and solvent preheating is supplied by burning heating oil.
- (13) The operating factor of the geothermal power plant is assumed as 75 percent.

The capital investment and annual operating cost are summarized in Tables 10 and 11 respectively and details of the calculation for these costs are shown in Appendix IV.

As can be seen from Table 10, the total capital investment for the steam treatment plant is \$3,118,000. The portion of the total capital investment is the reactor cost which is 35.28 percent of the total cost. The second largest capital item is the cost of offsite facilities (31.43%). The annual operating cost is \$416,750/year which is translated into 1.27 mil/kWh. The percentage distribution of the annual operating cost is: depreciation 37.41%, labor 29.51%, power 1.24%, fuel 26.51% and make-up solvent 5.33%. These figures show that the equipment cost is an important cost item in the overall steam treatment cost, and any reduction in the cost of equipment by improvement of process efficiency will strongly affect the cost of steam treatment.

It is probable that recovered sulfur would be a marketable commodity. Assuming that it would be salable at the present market level (~\$60/ton), a credit of about \$40,000 per year could be generated if the raw steam contains 200 ppm of H<sub>2</sub>S. This would reduce the operating cost by about nine percent.

TABLE 10. Capital Investment for Steam Treatment Plant  
(50 MWe Geothermal Power Plant)

1. Carbon	\$ 26,000
2. Solvent	70,000
3. Reactors	1,100,000
4. Solvent Storage Tanks	252,000
5. Air Compressor	40,000
6. Heat Exchangers	60,000
7. Crystallizer	40,000
8. Centrifuge	40,000
9. Superheaters (furnaces)	245,000
10. Instrumentation, 5% of (3+4+5+6+7+8+9)	85,000
11. Piping & Valves, 10% of (3+4+5+6+7+8+9+10)	180,000
12. Offsite Facilities, 50% of (3+4+5+6+7+8+9+10+11)	<u>980,000</u>
Total Capital Investment	\$3,118,000

TABLE 11. Annual Operating Cost for Steam Treatment Plant  
(50 MWe Geothermal Power Plant)

	<u>\$/Year</u>
14. Depreciation, 5% of (13)	155,900
15. Labor Cost	
4 operators   19,500x4 = 78,000	
1 maintenance           21,500	
1 supervisor <u>23,500</u>	
	123,000
16. Power (including offsite and auxiliary equipment)	5,150
17. Fuel	110,500
18. Make-up Solvent	<u>12,200</u>
19. Total Annual Operating Cost (14+15+16+17+18)	416,750/year
Steam Treatment Cost	1.27 mil/kWh

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APPENDIX I

EXPERIMENTAL BENCH SCALE SYSTEM

APPENDIX I  
EXPERIMENTAL BENCH SCALE SYSTEM

EQUIPMENT SETUP

To perform experiments under geothermal steam conditions, a bench scale gas removal testing apparatus has been constructed. A schematic flow diagram of this apparatus is shown in Figure I-1.

Water is charged to the steam generator<sup>(a)</sup> by a positive-displacement pump.<sup>(b)</sup> The boiler is operated at 180 psig. The steam generated passes through a steam superheater where it is heated to the desired condition. The steam superheater is a one-inch schedule 80, 11-inch long stainless steel pipe. The pipe is packed with aluminum pellets (3/16" x 3/16"), and its outer surface is coiled with a 76-inch tubular electrical heater.<sup>(c)</sup> The temperature of the superheated steam is controlled by a Variac connected to the heater.

The superheated steam flows to the reactor through a 1/2-inch stainless steel pipe traced with heating tape and insulated with Kaowool. Before entering the reactor, steam is mixed with noncondensable gases coming from the gas mixing station. Various combinations and compositions of noncondensable gases such as H<sub>2</sub>S, CO<sub>2</sub>, CH<sub>4</sub>, H<sub>2</sub>, NH<sub>3</sub>, and N<sub>2</sub> can be prepared and introduced to the steam through the gas mixing station. Also, air or oxygen, through a flowmeter, can be independently charged to the reactor system. The packed bed reactor is a one-inch schedule 80 stainless steel pipe 24 inches long with a support to hold the catalyst in the reactor. A detailed design of the reactor is shown in Figure I-2. A two-inch reactor was also constructed, but not used in this work.

The treated steam-gas mixture leaves the reactor and passes through the back pressure regulator,<sup>(d)</sup> which maintains the experimental unit at the

- 
- (a) Chromalox - Electric Steam Boiler (CHPE-18)
  - (b) Milroyal, Milton Roy Co.
  - (c) Chromalox TRL 7612
  - (d) Type 123 Diaphragm Operated Relief Valve

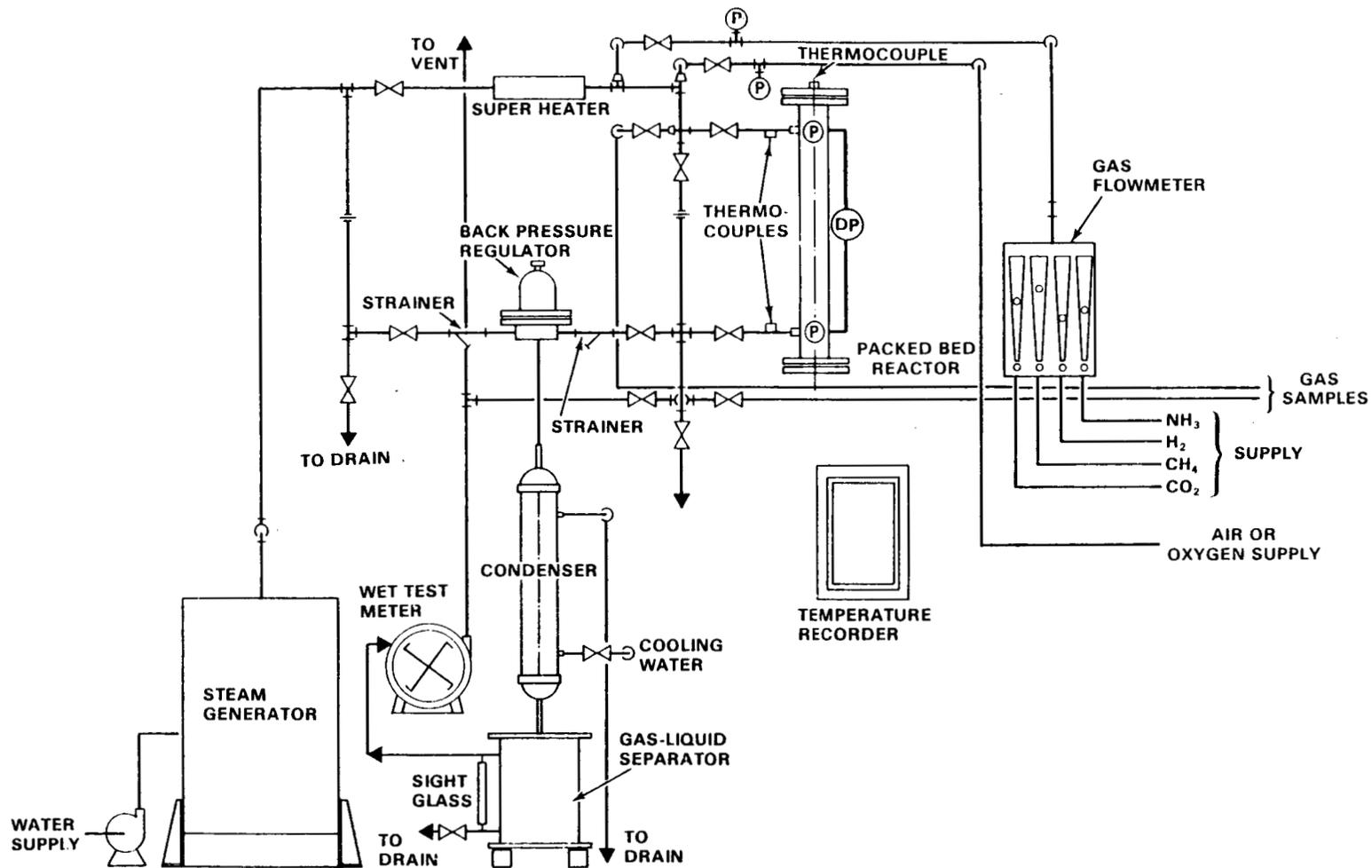


FIGURE I-1. Bench Scale H<sub>2</sub>S Removal Testing Unit

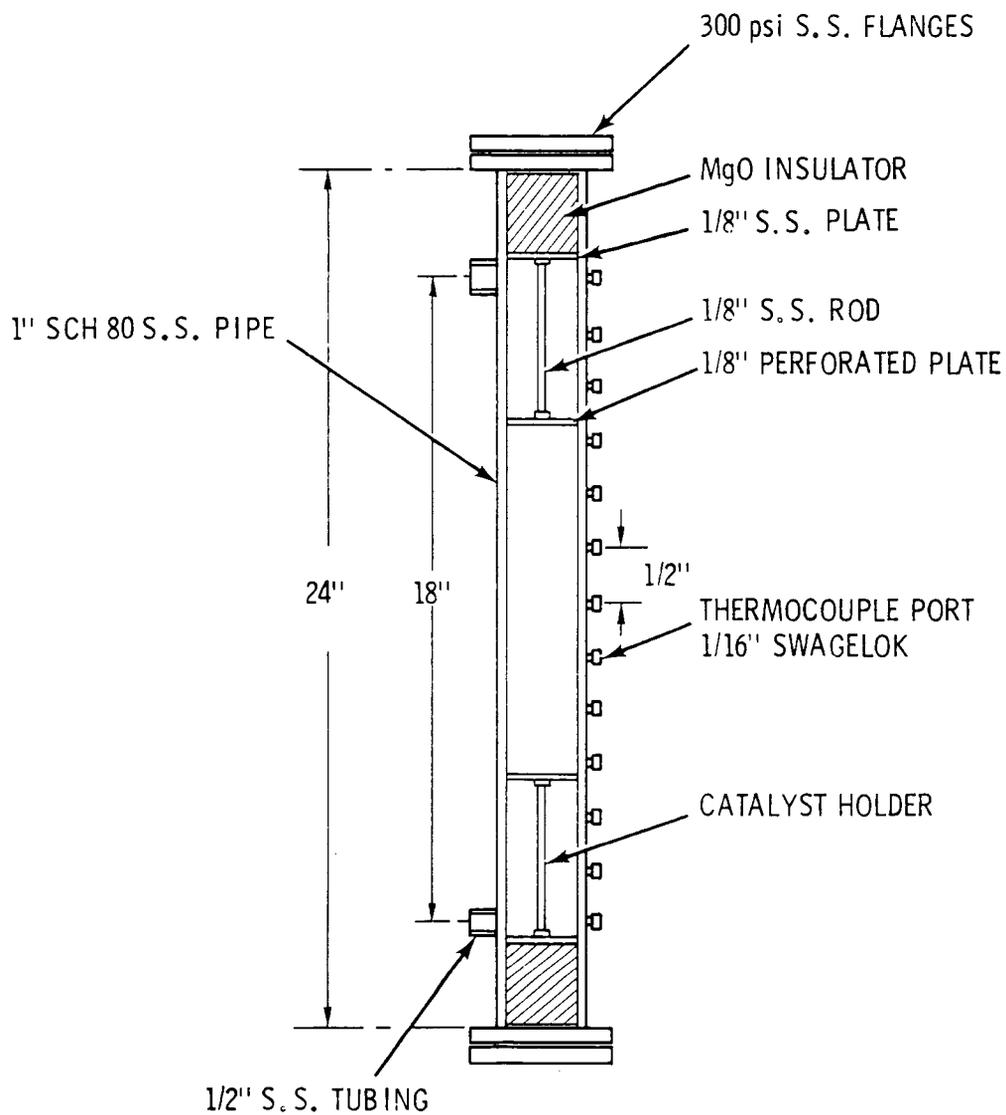


FIGURE I-2. Detailed Design of Reactor

desired pressure (nominally 100 psig). Steam from the regulator, after being reduced to one atmospheric pressure, is condensed, and the condensate is collected in the gas-liquid separator. Here the unreacted noncondensable gases such as  $N_2$ ,  $CO_2$ ,  $CH_4$ ,  $H_2$  and  $O_2$  are separated from the condensate and flow to a wet test meter where the total volume of unreacted noncondensable gases is measured. The effluent of the wet test meter is vented to the atmosphere through the building ventilation system. Gas samples are taken from two places; one at the inlet and one at the outlet of the reactor.

### GAS ANALYSIS

A wet chemical analysis procedure was developed in our laboratory based on an ASTM<sup>(10)</sup> method. Reagents used in this method include: a two percent zinc acetate solution, 0.1 N standard iodine and thiosulfate solutions, concentrated hydrochloric acid and a starch solution indicator. A 250 ml Erlenmeyer flask containing about 40 ml of zinc acetate solution is stored in a refrigerator to maintain the temperature of the solution as low as possible (without freezing). Prior to the collection of a gas sample, the flask with chilled zinc acetate solution is weighed. The mouth of the flask is plugged by a rubber stopper with a three-way valve. The flask is then connected to a vacuum system to evacuate the air in the flask. The air-free flask containing zinc acetate solution is immersed in an ice bath. To collect a sample, the flask is connected to one of the sampling lines shown in Figure I-1, and steam is vented through the three-way valve to the atmosphere for a few seconds. The three-way valve is then switched to a position where the steam sample flows into the flask. Due to the temperature of the zinc acetate solution and the ice bath, the condensation of steam takes place in the flask. When a suitable amount of sample is collected depending upon the concentration of  $H_2S$  in the sample, the three-way valve is switched to a new position where gas will not escape from the flask. The flask is then disconnected from the sampling line, and it is vigorously shaken for a few minutes to capture all the  $H_2S$  in the flask in the zinc acetate solution.  $H_2S$  reacts with the solution to form white zinc sulfide precipitate. After the reaction is completed, the flask is weighed again. The weight difference between the

initial and final weight is designated as the weight of sample. Here, we assume that the weight of the  $H_2S$  that reacted and any other noncondensable gases dissolved in the solution is negligibly small compared with the weight of steam condensate in the flask. The amount of  $H_2S$  that reacted with zinc acetate is then determined by an iodometric titration technique. In the titration a known amount of acidified (with concentrated hydrochloric acid) standard iodine solution is poured into the Erlenmeyer flask containing the gas sample. The hydrochloric acid reacts with zinc sulfide to release  $H_2S$  which immediately reacts with iodine solution according to the following reaction:



The excess iodine is back titrated with the standard thiosulfate solution using the starch solution as an indicator. The amount of  $H_2S$  in the sample can then be calculated from the actual amount of iodine consumed in Reaction (A1-1). This  $H_2S$  analysis method not only can be applied to the inlet and outlet streams of the reactor, but also it can be used to determine the dissolved  $H_2S$  in the steam condensate collected in the liquid-gas separator.

APPENDIX II

DETAILED EXPERIMENTAL DATA

Run#: 91, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 211.20 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
1.00	188	100*	29.09	86.23
1.83	188	100	75.58	64.21
2.33	188	100	63.90	69.74
4.66	188	100	42.40	79.92

Note: \* Saturated steam temperature is 170°C

Run#: 93, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 176.20 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.56	170	100*	133.03	24.50
1.95	170	100	143.70	18.44
2.37	170	100	144.06	18.24

Note: \* Saturated steam temperature is 170°C

I-II

Run#: 92, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 184.43 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.55	198	100*	4.30	97.67
1.00	198	100	5.68	96.92
2.00	198	100	11.40	93.82
3.00	198	100	16.40	91.11
4.00	198	100	21.51	88.34

Note: \* Saturated steam temperature is 170°C

Run#: 94, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 195.90 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.13	175	100*	4.22	97.85
1.17	175	100	23.30	88.11
2.05	175	100	19.70	89.94
2.80	180	100	27.10	86.17
3.05	171	100	78.70	59.83

Note: \* Saturated steam temperature is 170°C

Run#: 95, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 148.82 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.17	240	100*	0.00	100
1.92	250	100	25.60	82.80
2.87	258	100	13.70	90.79
3.92	258	100	12.35	91.70

Note: \* Saturated steam temperature is 170°C

Run#: 98, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 227.82 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
1.17	210	80*	3.93	98.27
1.83	210	80	17.23	92.44
3.25	190	80	9.20	95.96
4.25	185	80	27.50	87.93
5.67	162	80	35.87	84.25

Note: \* Saturated steam temperature is 162°C

II-2

Run#: 96, Catalyst: Sample 1\*  
 Average Inlet H<sub>2</sub>S Concentration: \_\_\_\_\_ ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
	182	100	80**	
	200	100	112	
	226	100	96	
	235	100	132	
	245	100	168	
	252	100	128	
	261	100	264	
	280	100	468	

Note: \*Sulfur saturated carbon was purged with superheated steam (44 lb/ft<sup>2</sup>/min)

\*\* Entrained sulfur concentration in steam

Regenerated catalyst from  
 Run#: 99, Catalyst: Run# 98  
 Average Inlet H<sub>2</sub>S Concentration: 197.30 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.28	151	80*	228.8	0.0
0.92	159	80	234.6	0.0
1.17	180	77	5.9	97.01
1.92	189	80	11.7	94.01
2.42	158	80	73.9	62.54

Note: \* Saturated steam temperature is 162°C

Run#: 100, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 254.8 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.37	175	80*	60.5	76.26
0.78	175	80	6.4	97.49
1.12	175	80	84.5	66.84
1.62	175	80	1.3	99.49
1.95	175	80	2.4	99.06
2.62	175	80	8.4	96.70
2.95	175	80	81.57	67.99
3.20	175	80	2.42	99.05

Note: \* Saturated steam temperature is 162°C

Run#: 104, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 262.2 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.28	173	120*	177.2	32.42
0.78	172	120	189.5	27.73
1.65	200	120	15.5	94.09
1.95	185	120	19.9	92.41
2.37	204	120	23.3	91.11
2.78	235	120	32.2	87.72

Note: \* Saturated steam temperature is 177°C

Run#: 103, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 213.85 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
1.08	140	38-27*	154	27.99
2.50	153	38-27	7.57	96.46
2.75	153	38-27	20.40	90.46
3.58	173	38-27	39.79	81.42
4.50	176	38-27	66.7	68.81
5.42	188	38-27	48.2	77.46

Note: \* Saturated steam temperature is 140°C @ 40psig  
 Pressure of the system was unstable in this range

Run#: 106, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 216.5 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
1.33	198	120*	7.8	96.40
1.75	184	120	29.0	86.37
2.03	182	120	33.9	84.34
4.25	197	120	56.2	74.04
5.08	197	120	21.0	90.30
5.75	197	120	37.6	82.63
6.75	182	120	67.9	68.64

Note: \* Saturated steam temperature is 177°C

Run#: 108, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 214.5 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.67	170	110*	163.2	23.92
2.50	198	110	18.4	91.42
3.17	205	110	11.2	94.79
3.50	215	110	20.9	90.26
4.33	225	110	51.7	75.90
5.00	240	110	59.6	72.21
5.42	232	110	68.8	67.93
5.67	225	110	62.3	70.96

Note: \* Saturated steam temperature is 173°C

Run#: \_\_\_\_\_, Catalyst: \_\_\_\_\_  
 Average Inlet H<sub>2</sub>S Concentration: \_\_\_\_\_ ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal

Note:

II-4

Run#: 109, Catalyst: Sample 1  
 Average Inlet H<sub>2</sub>S Concentration: 231.68 ppm

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal
0.5	170	110*	199.5	13.89
2.42	190	110	18.75	91.91

Note: \* Saturated steam temperature is 173°C

Run#: \_\_\_\_\_, Catalyst: \_\_\_\_\_  
 Average Inlet H<sub>2</sub>S Concentration: \_\_\_\_\_ ppm.

Run Time Hr.	Temperature °C	Pressure psig	Outlet H <sub>2</sub> S Conc. ppm	% H <sub>2</sub> S Removal

Note:

### DATA SHEET FOR PRESSURE DROP EXPERIMENT

Date: 3/20/18, Fluid: Stearine, Bed Diameter: 2.43 cm, Bed Height: 28.26 cm

Material in Bed: Sample 5, Partical Size: 0.69 mm, Voidage fraction: 0.41

Temp. °C	162 <sup>(1)</sup>				166				170 <sup>(2)</sup>				177 <sup>(3)</sup>				182				188			
Pressure psig	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Flow Rate lb/ft <sup>2</sup> /min	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft	2.51	3.68	7.41	14.31	1.99	2.61	5.71	9.96	1.99	2.61	5.71	9.96	1.99	2.61	5.71	9.96	1.99	2.61	5.71	9.96	1.99	2.61	5.71	9.96
Pressure psig									100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Flow Rate lb/ft <sup>2</sup> /min									0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft									1.93	3.19	5.74	13.25	1.64	2.71	5.22	8.41	1.64	2.71	5.22	8.41	1.64	2.71	5.22	8.41
Pressure psig													120	120	120	120	120	120	120	120	120	120	120	120
Flow Rate lb/ft <sup>2</sup> /min													0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft													1.51	2.71	6.48	10.27	1.64	2.15	5.13	7.68	1.64	2.12	5.13	7.68

- Note: (1) Saturated pressure: 80 psig  
 (2) Saturated pressure: 100 psig  
 (3) Saturated pressure: 120 psig

### DATA SHEET FOR PRESSURE DROP EXPERIMENT

Date: 3/23/78, Fluid: Steam, Bed Diameter: 2.43 cm, Bed Height: 29.69 cm

Material in Bed: Sample 2, Partical Size: 2.02 mm, Voidage fraction: 0.46

Temp. °C	162 <sup>(1)</sup>				166				170 <sup>(2)</sup>				177 <sup>(3)</sup>				182				188							
Pressure psig	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Flow Rate lb/ft <sup>2</sup> /min	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δp psi/ft	0.70	1.09	1.99	2.68	0.44	0.60	1.19	2.28	0.44	0.60	1.19	2.28	0.44	0.60	1.19	2.68	0.44	0.60	1.19	2.28	0.44	0.60	1.19	2.28	0.44	0.60	1.19	2.28
Pressure psig									100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Flow Rate lb/ft <sup>2</sup> /min									0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δp psi/ft									0.70	0.99	1.58	2.28	0.44	0.60	1.14	1.78	0.44	0.60	1.14	1.79	0.44	0.60	1.14	1.79	0.44	0.60	1.14	1.79
Pressure psig													120	120	120	120	120	120	120	120	120	120	120	120	120	120	120	120
Flow Rate lb/ft <sup>2</sup> /min													0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δp psi/ft													0.70	1.09	1.59	2.18	0.44	0.60	1.09	1.49	0.44	0.60	1.09	1.49	0.44	0.60	1.09	1.49

Note: (1) Saturated pressure: 80 psig  
 (2) Saturated pressure: 100 psig  
 (3) Saturated pressure: 120 psig

### DATA SHEET FOR PRESSURE DROP EXPERIMENT

Date: 3/21/78, Fluid: Steam, Bed Diameter: 2.43 cm, Bed Height: 29.69 cm

Material in Bed: Glass Beads, Partical Size: 6 mm, Voidage fraction: 0.435

Temp. °C	162 <sup>(1)</sup>				166				170 <sup>(2)</sup>				177 <sup>(3)</sup>				182				188							
Pressure psig	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Flow Rate lb/ft <sup>2</sup> /min	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.54	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft	0.31	0.49	0.63	1.18	0.07	0.11	0.60	0.77	0.07	0.11	0.56	0.77	0.07	0.11	0.60	0.77	0.07	0.07	0.60	0.70	0.07	0.22	0.56	0.90	0.07	0.22	0.56	0.90
Pressure psig									100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Flow Rate lb/ft <sup>2</sup> /min									0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft									0.21	0.31	0.63	0.77	0.07	0.17	0.28	0.70	0.07	0.17	0.28	0.70	0.07	0.17	0.28	0.70	0.07	0.17	0.28	0.70
Pressure psig													120	120	120	120	120	120	120	120	120	120	120	120	120	120	120	120
Flow Rate lb/ft <sup>2</sup> /min													0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft													0.31	0.35	0.49	0.77	0.18	0.21	0.49	0.70	0.14	0.31	0.45	0.63	0.14	0.31	0.45	0.63

Note: (1) Saturated pressure: 80 psig  
 (2) Saturated pressure: 100 psig  
 (3) Saturated pressure: 120 psig

### DATA SHEET FOR PRESSURE DROP EXPERIMENT

Date: 3/27/78, Fluid: Steam, Bed Diameter: 2.43 cm, Bed Height: 31.12 cm

Material in Bed: Glass Beads, Partical Size: 3 mm, Voidage fraction: 0.37

Temp. °C	162 <sup>(1)</sup>				166				170 <sup>(2)</sup>				177 <sup>(3)</sup>				182				188							
Pressure psig	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Flow Rate lb/ft <sup>2</sup> /min	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft	1.14	1.71	2.84	3.69	0.66	0.80	1.61	3.03	0.66	0.80	1.61	3.03	0.66	0.80	1.61	3.03	0.66	0.80	1.61	3.03	0.66	0.80	1.61	3.03	0.66	0.80	1.61	3.03
Pressure psig									100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Flow Rate lb/ft <sup>2</sup> /min									0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft									0.66	1.23	2.31	3.41	0.47	0.38	1.23	2.37	0.47	0.38	1.23	2.37	0.47	0.38	1.23	2.37	0.47	0.38	1.23	2.37
Pressure psig													120	120	120	120	120	120	120	120	120	120	120	120	120	120	120	120
Flow Rate lb/ft <sup>2</sup> /min													0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ p psi/ft													0.66	1.32	2.31	3.13	0.28	0.47	1.04	1.61	0.28	0.47	1.04	1.61	0.28	0.47	1.04	1.61

Note: (1) Saturated pressure: 80 psig  
 (2) Saturated pressure: 100 psig  
 (3) Saturated pressure: 120 psig

### DATA SHEET FOR PRESSURE DROP EXPERIMENT

Date: 3/31/78, Fluid: Stearine, Bed Diameter: 2.43 cm, Bed Height: 30.02 cm

Material in Bed: Sample 6, Partical Size: 0.53 mm, Voidage fraction: 0.43

Temp. °C	162 <sup>(1)</sup>				166				172 <sup>(2)</sup>				177 <sup>(3)</sup>				182				188							
Pressure psig	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80	80
Flow Rate lb/ft <sup>2</sup> /min	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft	7.25	11.32	16.63	-	3.87	6.00	11.97	17.54	3.87	6.00	11.97	-	3.87	6.01	12.07	20.12	3.87	6.09	12.19	20.12	3.87	6.09	12.57	20.31	3.87	6.09	12.57	20.31
Pressure psig									100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Flow Rate lb/ft <sup>2</sup> /min									0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft									5.51	8.90	16.15	23.77	3.58	5.13	9.57	15.38	3.58	5.13	9.57	15.38	3.58	5.13	9.57	15.38	3.58	5.13	9.57	15.38
Pressure psig													120	120	120	120	120	120	120	120	120	120	120	120	120	120	120	120
Flow Rate lb/ft <sup>2</sup> /min													0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47	0.59	0.74	1.11	1.47
Δ P psi/ft													4.84	7.35	14.80	23.01	3.09	3.87	5.03	9.71	3.09	3.87	5.03	9.67	3.09	3.87	5.03	9.67

Note: (1) Saturated pressure: 80 psig  
 (2) Saturated pressure: 100 psig  
 (3) Saturated pressure: 120 psig

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# R1-2, Date: 8/21/78

Purpose of Run: The spent carbon of Run R1-1 was steam purged to remove sulfur. The resulting regenerated carbon was used in this run.

Catalyst: Reg. of Run R1-1  
Sample 5, Weight: 80 g, Bed Height: 26.83 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 192 °C, Outlet: 192 °C

Flow Rate: 25.47 g/cm<sup>2</sup>/min., 52.16 lb/ft<sup>2</sup>/min.

Space Velocity: 221.97 v/v/min., Residence Time: 0.2703 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 1749, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.57

Total Run Time: 5.16 hr

Sulfur Balance: Charged to System: 0.1729 g moles

On Catalyst: 0.1661 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: \_\_\_\_\_ g moles

Overall Sulfur Recovered: 96.09 %\*, ( \_\_\_\_\_ %) \*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 6.83 ppm, O<sub>2</sub>: \_\_\_\_\_ ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.5</u>	<u>0</u>				
<u>1.0</u>	<u>3.08</u>				
<u>3.0</u>	<u>5.07</u>				
<u>4.0</u>	<u>68.60</u>				
<u>5.16</u>	<u>21.38</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# R1-1, Date: 8/17/78

Purpose of Run: Spent carbon (Samples) after CS<sub>2</sub> extraction and steam treatment was used to test the activity of regenerated cat.

Catalyst: Reg. Sample 5, Weight: 80 g, Bed Height: 26.83 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 189 °C, Outlet: 189 °C

Flow Rate: 23.27 g/cm<sup>2</sup>/min., 47.66 lb/ft<sup>2</sup>/min.

Space Velocity: 220.90 v/v/min., Residence Time: 0.2716 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 1979, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.39

Total Run Time: 14.8 hr

Sulfur Balance: Charged to System: 0.5587 g moles

On Catalyst: 0.5414 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0083 g moles

Overall Sulfur Recovered: 96.90 %\*, ( \_\_\_\_\_ %) \*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 6.13 ppm, O<sub>2</sub>: 39.30 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.92</u>	<u>2.92</u>	<u>9.58</u>	<u>6.42</u>	<u>14.17</u>	<u>16.62</u>
<u>2.00</u>	<u>1.07</u>	<u>10.08</u>	<u>6.47</u>	<u>14.75</u>	<u>17.35</u>
<u>3.00</u>	<u>2.08</u>	<u>10.83</u>	<u>10.70</u>		
<u>4.25</u>	<u>3.74</u>	<u>11.33</u>	<u>10.20</u>		
<u>5.33</u>	<u>3.09</u>	<u>11.83</u>	<u>10.33</u>		
<u>6.92</u>	<u>3.80</u>	<u>12.33</u>	<u>15.24</u>		
<u>7.83</u>	<u>4.62</u>	<u>12.83</u>	<u>14.66</u>		
<u>8.83</u>	<u>4.88</u>	<u>13.50</u>	<u>16.27</u>		

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 163-1, Date: 8/29/78

Purpose of Run: Carbon regenerated with CS<sub>2</sub>, then treated with NH<sub>3</sub>  
Regenerated from Run 163

Catalyst: Sample 1, Weight: 80 g, Bed Height: 29.21 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 183 °C, Outlet: 189 °C

Flow Rate: 23.61 g/cm<sup>2</sup>/min., 48.35 lb/ft<sup>2</sup>/min.

Space Velocity: 206.05 v/v/min., Residence Time: 0.2912 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 208.8 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.55

Total Run Time: 15 hr

Sulfur Balance: Charged to System: 0.6032 g moles

On Catalyst: 0.5860 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0155 g moles

Overall Sulfur Recovered: 97.15 %\*, ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 5.94 ppm, O<sub>2</sub>: 51.91 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.83</u>	<u>2.41</u>	<u>10.67</u>	<u>10.92</u>		
<u>1.42</u>	<u>0.77</u>	<u>11.67</u>	<u>11.05</u>		
<u>4.08</u>	<u>1.92</u>	<u>12.67</u>	<u>14.89</u>		
<u>6.85</u>	<u>1.39</u>	<u>13.75</u>	<u>18.16</u>		
<u>7.85</u>	<u>11.57</u>	<u>14.92</u>	<u>17.38</u>		
<u>8.50</u>	<u>8.87</u>				
<u>9.42</u>	<u>8.92</u>				
<u>10.17</u>	<u>14.87</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 163, Date: 8/11/78

Purpose of Run: Effect of noncondensable gas on H<sub>2</sub>S Removal Capacity

Catalyst: Sample 1, Weight: 80 g, Bed Height: 36.83 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 184 °C, Outlet: 193 °C

Flow Rate: 22.66 g/cm<sup>2</sup>/min., 46.41 lb/ft<sup>2</sup>/min.

Space Velocity: 156.69 v/v/min., Residence Time: 0.38 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 193, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: \_\_\_\_\_

Total Run Time: 27.08 hr

Sulfur Balance: Charged to System: 0.9707 g moles

On Catalyst: 0.9449 g moles\*, (0.7622 g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.02 g moles

Overall Sulfur Recovered: 97.35 %\*, (78.52 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 5.91 ppm, O<sub>2</sub>: 42.48 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.67</u>	<u>0.75</u>	<u>15.00</u>	<u>2.42</u>	<u>21.33</u>	<u>7.22</u>
<u>2.83</u>	<u>0.92</u>	<u>15.75</u>	<u>4.99</u>	<u>22.25</u>	<u>9.68</u>
<u>5.08</u>	<u>1.06</u>	<u>16.50</u>	<u>5.07</u>	<u>24.00</u>	<u>11.55</u>
<u>7.17</u>	<u>1.51</u>	<u>17.50</u>	<u>4.74</u>	<u>25.00</u>	<u>16.94</u>
<u>8.42</u>	<u>1.94</u>	<u>18.25</u>	<u>8.82</u>	<u>26.00</u>	<u>17.21</u>
<u>9.75</u>	<u>1.25</u>	<u>19.00</u>	<u>10.29</u>	<u>27.00</u>	<u>17.62</u>
<u>10.17</u>	<u>1.70</u>	<u>19.75</u>	<u>7.53</u>		
<u>12.75</u>	<u>1.49</u>	<u>20.50</u>	<u>6.83</u>		

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 159-3, Date: 9/27/78

Purpose of Run: Activity of Carbon After three times of dichloroethane regeneration  
regenerated from Run 159-2

Catalyst: Sample, Weight: 70.70 g, Bed Height: 30.48 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 188 °C, Outlet: 184 °C

Flow Rate: 19.99 g/cm<sup>2</sup>/min., 40.22 lb/ft<sup>2</sup>/min.

Space Velocity: 164.25 v/v/min., Residence Time: 0.3653 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 2013, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.63

Total Run Time: 13 hr

Sulfur Balance: Charged to System: 0.4204 g moles

On Catalyst: 0.4188 g moles\* ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery Dichloroethane extraction

In Steam Condensate: 99.63 g moles

Overall Sulfur Recovered: 99.63%\* ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 0.77 ppm, O<sub>2</sub>: 59.18 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.25</u>	<u>4.70</u>	<u>13.00</u>	<u>9.71</u>		
<u>3.33</u>	<u>3.34</u>				
<u>5.42</u>	<u>0.</u>				
<u>6.75</u>	<u>0.</u>				
<u>8.42</u>	<u>0.</u>				
<u>9.92</u>	<u>1.08</u>				
<u>11.50</u>	<u>3.47</u>				
<u>12.58</u>	<u>5.92</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 159-2, Data: 9/22/78

Purpose of Run: Activity of Carbon After two times of dichloroethane regeneration  
Regenerated from Run 159-1

Catalyst: Sample 5, Weight: 80 g, Bed Height: 27.31 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 185 °C, Outlet: 181 °C

Flow Rate: 20.13 g/cm<sup>2</sup>/min., 41.23 lb/ft<sup>2</sup>/min.

Space Velocity: 187.98 v/v/min., Residence Time: 0.3192 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 2073, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.56

Total Run Time: 12.5 hr

Sulfur Balance: Charged to System: 0.4268 g moles

On Catalyst: 0.4183 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery Dichloroethane extraction

In Steam Condensate: 0.0183 g moles

Overall Sulfur Recovered: 98.00%, ( \_\_\_\_\_ )\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 4.13 ppm, O<sub>2</sub>: 53.17 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.08</u>	<u>10.94</u>				
<u>2.83</u>	<u>4.49</u>				
<u>4.92</u>	<u>4.22</u>				
<u>5.50</u>	<u>2.52</u>				
<u>9.19</u>	<u>5.34</u>				
<u>10.33</u>	<u>8.12</u>				
<u>11.42</u>	<u>12.29</u>				
<u>12.42</u>	<u>16.80</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 159-1, Data: 9/8/78

Purpose of Run: Activity of dichloroethane regenerated carbon

Catalyst: Regenerated Sample 5, Weight: 76.60 g, Bed Height: 25.40 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 190 °C, Outlet: 195 °C

Flow Rate: 21.57 g/cm<sup>2</sup>/min., 44.18 lb/ft<sup>2</sup>/min.

Space Velocity: 216.5 v/v/min., Residence Time: 0.2771 sec.

Noncondensable Gas: H<sub>2</sub>S, NH<sub>3</sub>, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, N<sub>2</sub>

Concentration, ppm: 188.4 55, 6210, 52, 385, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.6311

Total Run Time: 13.75 hr

Sulfur Balance: Charged to System: 0.4572 g moles

On Catalyst: 0.4419 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery Dichloroethane extraction

In Steam Condensate: \_\_\_\_\_ g moles

Overall Sulfur Recovered: 96.65%\*, ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 6.31 ppm, O<sub>2</sub>: 54.07 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.09</u>	<u>1.15</u>	<u>10.17</u>	<u>4.80</u>		
<u>2.92</u>	<u>2.10</u>	<u>11.08</u>	<u>6.35</u>		
<u>3.75</u>	<u>2.70</u>	<u>11.75</u>	<u>7.19</u>		
<u>5.08</u>	<u>5.59</u>	<u>12.67</u>	<u>13.8</u>		
<u>6.08</u>	<u>2.13</u>	<u>13.25</u>	<u>16.9</u>		
<u>7.08</u>	<u>2.78</u>	<u>13.75</u>	<u>23.2</u>		
<u>8.83</u>	<u>2.37</u>				
<u>9.33</u>	<u>2.82</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 159, Date: 8/3/78

Purpose of Run: Variation of Oxygen supply rate

Catalyst: Sample 5, Weight: 80 g, Bed Height: 25.4 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 185 °C, Outlet: 191 °C

Flow Rate: 21.30 g/cm<sup>2</sup>/min., 43.63 lb/ft<sup>2</sup>/min.

Space Velocity: 213.85 v/v/min., Residence Time: 0.2806 sec.

Noncondensable Gas: H<sub>2</sub>S, NH<sub>3</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 2203, 55, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.14

Total Run Time: 6.75 hr

Sulfur Balance: Charged to System: 0.2592 g moles

On Catalyst: 0.2486 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0011 g moles

Overall Sulfur Recovered: 95.94%,\* ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 9.01 ppm, O<sub>2</sub>: 13.76 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, -hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.75</u>	<u>0.03</u>				
<u>1.58</u>	<u>1.00</u>				
<u>3.92</u>	<u>0.94</u>				
<u>5.42</u>	<u>2.93</u>				
<u>6.00</u>	<u>3.93</u>				
<u>6.50</u>	<u>3.95</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 152, Data: 7/12/78

Purpose of Run: Effect of noncondensable gas on H<sub>2</sub>S Removal Capacity

Catalyst: Sample 1, Weight: 80 g, Bed Height: 36.20 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 186 °C, Outlet: 191 °C

Flow Rate: 18.64 g/cm<sup>2</sup>/min., 38.18 lb/ft<sup>2</sup>/min.

Space Velocity: 131.38 v/v/min., Residence Time: 0.46 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 2165, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.59

Total Run Time: 14.0 hr

Sulfur Balance: Charged to System: 0.4622 g moles

On Catalyst: 0.4565 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0019 g moles

Overall Sulfur Recovered: 98.79%\*, ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 2.67 ppm, O<sub>2</sub>: 61.09 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.92</u>	<u>2.52</u>	<u>12.00</u>	<u>1.00</u>		
<u>2.92</u>	<u>1.36</u>	<u>13.08</u>	<u>1.64</u>		
<u>4.58</u>	<u>2.50</u>	<u>14.00</u>	<u>0.85</u>		
<u>5.67</u>	<u>1.22</u>				
<u>8.00</u>	<u>2.92</u>				
<u>8.83</u>	<u>0.92</u>				
<u>10.17</u>	<u>4.09</u>				
<u>11.08</u>	<u>1.00</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 151 , Data: 7/11/78

Purpose of Run: Effect of NH<sub>3</sub> on H<sub>2</sub>S removal Capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 25.72 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 186 °C, Outlet: 188 °C

Flow Rate: 18.87 g/cm<sup>2</sup>/min., 38.66 lb/ft<sup>2</sup>/min.

Space Velocity: 178.20 v/v/min., Residence Time: 0.3367 sec.

Noncondensable Gas: H<sub>2</sub>S, NH<sub>3</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 215, 55, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.58

Total Run Time: 10 hr

Sulfur Balance: Charged to System: 0.3320 g moles

On Catalyst: 0.3302 g moles\*, (0.1863 g moles)\*\*

Method of Sulfur Recovery CS<sub>2</sub> extraction

In Steam Condensate: 0.0007 g moles

Overall Sulfur Recovered: 99.44 %\*, (56.10 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.1657 ppm, O<sub>2</sub>: 59.1238 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>2.33</u>	<u>2.52</u>				
<u>5.33</u>	<u>0.94</u>				
<u>7.08</u>	<u>1.32</u>				
<u>10.00</u>	<u>2.37</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 150, Date: 7/10/78

Purpose of Run: Effect of all noncondensable gas (concentration same as that in Geysers) on H<sub>2</sub>S removal capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 25.72 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 186 °C, Outlet: 188 °C

Flow Rate: 18.69 g/cm<sup>2</sup>/min., 38.29 lb/ft<sup>2</sup>/min.

Space Velocity: 176.50 v/v/min., Residence Time: 0.34 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 2079, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.65

Total Run Time: 10 hr

Sulfur Balance: Charged to System: 0.3179 g moles

On Catalyst: 0.3161 g moles\*, (0.1500 g moles)\*\*

Method of Sulfur Recovery CS<sub>2</sub> extraction

In Steam Condensate: 0.0011 g moles

Overall Sulfur Recovered: 99.42%, (47.18%)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.18 ppm, O<sub>2</sub>: 64.06 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>4.25</u>	<u>2.2</u>				
<u>6.58</u>	<u>0.77</u>				
<u>8.75</u>	<u>1.21</u>				
<u>10:00</u>	<u>2.40</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 149, Date: 6/23/78

Purpose of Run: Effect of all noncondensable gas (High Conc.) on H<sub>2</sub>S removal capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 28.89 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 185 °C, Outlet: 187 °C

Flow Rate: 18.76 g/cm<sup>2</sup>/min., 38.43 lb/ft<sup>2</sup>/min.

Space Velocity: 165.56 v/v/min., Residence Time: 0.3624 sec.

Noncondensable Gas: H<sub>2</sub>S, CH<sub>4</sub>, H<sub>2</sub>, CO<sub>2</sub>, NH<sub>3</sub>, \_\_\_\_\_

Concentration, ppm: 1953, 770, 104, 12420, 110, \_\_\_\_\_

Oxygen Supply: Air ~~to oxygen~~, Stoichiometric: 1.75

Total Run Time: 10.5 hr

Sulfur Balance: Charged to System: 0.3148 g moles

On Catalyst: 0.3116 g moles\*, (0.1734 g moles)\*\*

Method of Sulfur Recovery CS<sub>2</sub> extraction

In Steam Condensate: 0.0013 g moles

Overall Sulfur Recovered: 98.98 %\*, (55.09 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.99 ppm, O<sub>2</sub>: 69.72 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, -hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.42</u>	<u>0.80</u>				
<u>2.00</u>	<u>0.70</u>				
<u>3.58</u>	<u>0.63</u>				
<u>5.75</u>	<u>3.37</u>				
<u>6.83</u>	<u>0.62</u>				
<u>7.83</u>	<u>2.07</u>				
<u>9.83</u>	<u>-</u>				
<u>10.25</u>	<u>0.73</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 145, Date: 6/1/78

Purpose of Run: Effect of high NH<sub>3</sub> concentration on H<sub>2</sub>S removal capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 28.10 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 186 °C, Outlet: 190 °C

Flow Rate: 18.97 g/cm<sup>2</sup>/min., 38.86 lb/ft<sup>2</sup>/min.

Space Velocity: 171.16 v/v/min., Residence Time: 0.3485 sec.

Noncondensable Gas: H<sub>2</sub>S, NH<sub>3</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 209.1, 110, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.61

Total Run Time: 9 hr

Sulfur Balance: Charged to System: 0.2922 g moles

On Catalyst: 0.2844 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0022 g moles

Overall Sulfur Recovered: 97.33 %\*, ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 5.58 ppm, O<sub>2</sub>: 63.06 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>3.50</u>	<u>9.70</u>				
<u>4.50</u>	<u>7.13</u>				
<u>5.92</u>	<u>4.88</u>				
<u>7.00</u>	<u>4.73</u>				
<u>8.00</u>	<u>6.55</u>				
<u>8.92</u>	<u>4.94</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 142, Data: 5/17/78

Purpose of Run: Effect of high CH<sub>4</sub> conc. on H<sub>2</sub>S removal capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 28.58 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 187 °C, Outlet: 191 °C

Flow Rate: 19.22 g/cm<sup>2</sup>/min., 39.37 lb/ft<sup>2</sup>/min.

Space Velocity: 171.49 v/v/min., Residence Time: 0.3499 sec.

Noncondensable Gas: H<sub>2</sub>S, CH<sub>4</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 200, 770, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/~~Oxygen~~, Stoichiometric: 1.68

Total Run Time: 13 hr

Sulfur Balance: Charged to System: 0.4088 g moles

On Catalyst: 0.4049 g moles\*, (0.1325 g moles)\*\*

Method of Sulfur Recovery acetone extraction

In Steam Condensate: 0.0014 g moles

Overall Sulfur Recovered: 99.05 %\*, (32.41 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.90 ppm, O<sub>2</sub>: 63.45 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.08</u>	<u>2.35</u>	<u>11.58</u>	<u>0.68</u>		
<u>2.25</u>	<u>1.46</u>	<u>12.58</u>	<u>2.39</u>		
<u>4.25</u>	<u>3.27</u>	<u>12.92</u>	<u>0.67</u>		
<u>6.08</u>	<u>2.14</u>				
<u>7.58</u>	<u>1.99</u>				
<u>8.58</u>	<u>2.59</u>				
<u>9.58</u>	<u>1.79</u>				
<u>10.58</u>	<u>1.85</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 140, Date: 5/10/78

Purpose of Run: Effect of high H<sub>2</sub> conc. on H<sub>2</sub>S removal capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 28.10 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 191 °C, Outlet: 193 °C

Flow Rate: 19.23 g/cm<sup>2</sup>/min., 39.38 lb/ft<sup>2</sup>/min.

Space Velocity: 174.46 v/v/min., Residence Time: 0.3439 sec.

Noncondensable Gas: H<sub>2</sub>S, H<sub>2</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 207, 104, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/~~Oxygen~~, Stoichiometric: 1.61

Total Run Time: 13.5 hr

Sulfur Balance: Charged to System: 0.4396 g moles

On Catalyst: 0.4359 g moles\*, (0.2484 g moles)\*\*

Method of Sulfur Recovery Tetrachloroethylene extraction

In Steam Condensate: 0.0029 g moles

Overall Sulfur Recovered: 99.15 %\*, (56.51 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.74 ppm, O<sub>2</sub>: 60.10 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.00</u>	<u>1.06</u>	<u>12.83</u>	<u>4.29</u>		
<u>2.85</u>	<u>0.91</u>	<u>13.42</u>	<u>5.32</u>		
<u>4.33</u>	<u>1.92</u>				
<u>6.42</u>	<u>1.19</u>				
<u>8.58</u>	<u>1.27</u>				
<u>11.08</u>	<u>2.53</u>				
<u>11.75</u>	<u>1.76</u>				
<u>12.33</u>	<u>3.11</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 138-1, Date: 8/30/78 - 9/1/78

Purpose of Run: Activity of tetrachloroethylene regenerated Carbon

Catalyst: Sample 5, Weight: 103.82 g, Bed Height: 29.53 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 189 °C, Outlet: 189 °C

Flow Rate: 22.26 g/cm<sup>2</sup>/min., 45.60 lb/ft<sup>2</sup>/min.

Space Velocity: 192.04 v/v/min., Residence Time: 0.3124 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 209.3, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.38

Total Run Time: 13 hr

Sulfur Balance: Charged to System: 0.4955 g moles

On Catalyst: 0.4858 g moles\*, ( \_\_\_\_\_ g moles)\*\*

Method of Sulfur Recovery \_\_\_\_\_

In Steam Condensate: 0.0161 g moles

Overall Sulfur Recovered: 98.04 %\*, ( \_\_\_\_\_ %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 4.10 ppm, O<sub>2</sub>: 38.88 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.17</u>	<u>0.65</u>	<u>11.83</u>	<u>11.21</u>		
<u>2.33</u>	<u>1.19</u>	<u>12.92</u>	<u>12.96</u>		
<u>4.58</u>	<u>1.95</u>				
<u>5.50</u>	<u>0.61</u>				
<u>7.33</u>	<u>1.63</u>				
<u>8.42</u>	<u>16.35</u>				
<u>9.75</u>	<u>9.58</u>				
<u>11.00</u>	<u>8.57</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 138, Date: 5/5/78

Purpose of Run: Effect of high CO<sub>2</sub> conc. on H<sub>2</sub>S removal Capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 28.56 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 186 °C, Outlet: 190 °C

Flow Rate: 20.11 g/cm<sup>2</sup>/min., 41.18 lb/ft<sup>2</sup>/min.

Space Velocity: 181.41 v/v/min., Residence Time: 0.331 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 2164 12400, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air ~~oxygen~~, Stoichiometric: 1.497

Total Run Time: 13.50 hr

Sulfur Balance: Charged to System: 0.4796 g moles

On Catalyst: 0.4728 g moles\*, (0.3075 g moles)\*\*

Method of Sulfur Recovery Tetrachloroethylene extraction

In Steam Condensate: 0.0039 g moles

Overall Sulfur Recovered: 98.58 %\*, (64.12 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 3.02 ppm, O<sub>2</sub>: 48.99 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.75</u>	<u>0</u>	<u>9.25</u>	<u>0.82</u>		
<u>1.83</u>	<u>0.76</u>	<u>10.25</u>	<u>2.15</u>		
<u>3.55</u>	<u>0.75</u>	<u>11.17</u>	<u>5.53</u>		
<u>5.08</u>	<u>0.74</u>	<u>11.67</u>	<u>7.32</u>		
<u>6.08</u>	<u>0.78</u>	<u>12.25</u>	<u>10.64</u>		
<u>7.00</u>	<u>2.23</u>	<u>12.75</u>	<u>14.32</u>		
<u>7.67</u>	<u>2.28</u>	<u>13.17</u>	<u>14.71</u>		
<u>8.17</u>	<u>0.94</u>	<u>13.42</u>	<u>14.77</u>		

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 137, Date: 5/4/78

Purpose of Run: Effect of CO<sub>2</sub> on H<sub>2</sub>S removal Capacity

Catalyst: Sample 5, Weight: 80 g, Bed Height: 27.94 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 188 °C, Outlet: 192 °C

Flow Rate: 20.62 g/cm<sup>2</sup>/min., 42.23 lb/ft<sup>2</sup>/min.

Space Velocity: 188.16 v/v/min., Residence Time: 0.3189 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 216, 6200, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air ~~to oxygen~~, Stoichiometric: 1.45

Total Run Time: 10 hr

Sulfur Balance: Charged to System: 0.3583 g moles

On Catalyst: 0.3563 g moles\*, (0.1988 g moles)\*\*

Method of Sulfur Recovery Tetrachloroethylene extraction

In Steam Condensate: 0.0035 g moles

Overall Sulfur Recovered: 99.44 %\*, (55.48 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 1.20 ppm, O<sub>2</sub>: 44.66 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>0.75</u>	<u>1.63</u>	<u>7.58</u>	<u>0</u>		
<u>1.75</u>	<u>0</u>	<u>9.00</u>	<u>1.41</u>		
<u>2.75</u>	<u>3.53</u>	<u>9.75</u>	<u>2.42</u>		
<u>3.50</u>	<u>1.86</u>				
<u>3.83</u>	<u>2.29</u>				
<u>4.58</u>	<u>0.98</u>				
<u>5.58</u>	<u>1.83</u>				
<u>6.58</u>	<u>1.76</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 113-1, Date: 8/28/78

Purpose of Run: Spent catalyst of Run 113 was extracted with CS<sub>2</sub> and treated with NH<sub>3</sub>. To study the activity of regenerated cat.

Catalyst: Reg. cat. Sample 1, Weight: 79.41 g, Bed Height: 36.20 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 183 °C, Outlet: 187 °C

Flow Rate: 22.72 g/cm<sup>2</sup>/min., 46.54 lb/ft<sup>2</sup>/min.

Space Velocity: 159.88 v/v/min., Residence Time: 0.3753 sec.

Noncondensable Gas: H<sub>2</sub>S, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, NH<sub>3</sub>, N<sub>2</sub>

Concentration, ppm: 191.7, 6210, 52, 385, 55, 91

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.47

Total Run Time: 12.5 hr

Sulfur Balance: Charged to System: 0.4455 g moles

On Catalyst: 0.4335 g moles\*, (0.3291 g moles)\*\*

Method of Sulfur Recovery (NH<sub>3</sub>)<sub>2</sub>S + H<sub>2</sub>O extraction

In Steam Condensate: 0.0107 g moles

Overall Sulfur Recovered: 97.32 %\*, (73.86 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 5.16 ppm, O<sub>2</sub>: 44.92 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>2.75</u>	<u>1.67</u>	<u>12.33</u>	<u>15.35</u>		
<u>4.16</u>	<u>2.36</u>				
<u>6.33</u>	<u>0.60</u>				
<u>7.33</u>	<u>6.07</u>				
<u>8.33</u>	<u>9.22</u>				
<u>9.42</u>	<u>5.76</u>				
<u>10.50</u>	<u>7.81</u>				
<u>11.66</u>	<u>16.71</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

Summary Data for H<sub>2</sub>S Removal from Geothermal Steam Experiment

Run# 113 , Data: 9/6/77

Purpose of Run: To test the activity of regenerate catalyst from Run 112

Catalyst: Reg. Cat. from Run 112, Weight: 80 g, Bed Height: 34.29 cm

Steam: Pressure: 100 psig, Bed Temp. Inlet: 189 °C, Outlet: 190 °C

Flow Rate: 21.57 g/cm<sup>2</sup>/min., 44.18 lb/ft<sup>2</sup>/min.

Space Velocity: 159.4 v/v/min., Residence Time: 0.38 sec.

Noncondensable Gas: H<sub>2</sub>S, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Concentration, ppm: 190, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_, \_\_\_\_\_

Oxygen Supply: Air/Oxygen, Stoichiometric: 1.28

Total Run Time: 6.08 hr

Sulfur Balance: Charged to System: 0.2525 g moles

On Catalyst: 0.2275 g moles\* (0.1458 g moles)\*\*

Method of Sulfur Recovery CS<sub>2</sub> extraction

In Steam Condensate: 0.0092 g moles

Overall Sulfur Recovered: 90 %\* (58 %)\*\*

Average Gas Concentration in Reactor Effluent:

H<sub>2</sub>S: 22.92 ppm, O<sub>2</sub>: 41.25 ppm

H<sub>2</sub>S Concentration vs. Time in Treated Steam

Run Time, .hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm	Run Time, hr	H <sub>2</sub> S, ppm
<u>1.33</u>	<u>4.40</u>	<u>6.08</u>	<u>52.76</u>		
<u>1.92</u>	<u>10.60</u>				
<u>2.42</u>	<u>17.99</u>				
<u>2.92</u>	<u>16.32</u>				
<u>3.50</u>	<u>27.62</u>				
<u>4.17</u>	<u>31.43</u>				
<u>4.83</u>	<u>37.51</u>				
<u>5.50</u>	<u>46.38</u>				

\* Calculated from H<sub>2</sub>S concentration vs. time curve

\*\* Experimental value

APPENDIX III

EVALUATION OF TECHNIQUES FOR AVOIDING  
CONDENSATE IN CATALYST BED

APPENDIX III  
EVALUATION OF TECHNIQUES FOR AVOIDING CONDENSATE IN CATALYST BED

It has been proven that the catalytic oxidation process removes  $H_2S$  from the simulated geothermal steam only when condensation is avoided in the catalyst bed. Two possibilities were evaluated for avoiding this difficulty: throttling and superheating. These are discussed in the following paragraphs from the standpoint of energy efficiency.

THROTTLING

In the throttling process steam is adiabatically throttled to lower pressure superheated steam. Although the throttling process changes saturated steam to a superheated condition, it also causes degradation of the work available from the steam. The available work relates directly to the amount of power which can be potentially produced in a steam turbine power generator. The degradation of the available work is of importance in the power generation, and it must be minimized. To calculate the amount of available work <sup>(17)</sup> and the loss of available work, the following equation is used:

$$W^0 = h_i - h_0 - T_0 (s_i - s_0)$$

where:  $W^0$ : available work, Btu/lb  
 $h_i$ : enthalpy of steam, Btu/lb  
 $h_0$ : enthalpy of steam at a reference state, Btu/lb  
 $T_0$ : temperature of reference state,  $^{\circ}R$   
 $s_i$ : entropy of steam, Btu/ $^{\circ}R$ /lb  
 $s_0$ : entropy of steam at a reference state, Btu/ $^{\circ}R$ /lb.

The reference state is chosen as 4" Hg, 126 $^{\circ}F$  saturated steam which roughly corresponds to the exit condition of the steam from turbine power generators of geothermal power plants operated at the Geysers.

The available work of the saturated steam,  $W_{sat}^0$ , can be calculated by substituting H and s values from a steam table.

Since the steam throttling process is an adiabatic process, enthalpies of steam before and after throttling are the same. Following the constant enthalpy line in Mollier's chart the properties of the superheated steam such as temperature, pressure, enthalpy and entropy can be obtained. The available work from the superheated steam  $W_{sup}^0$  can also be calculated from the above equation.

The loss of the available work,  $\Delta W^0$ , due to the throttling process equals the difference between  $W_{sat}^0$  and  $W_{sup}^0$ . The ratio R of available work of the superheated steam to that of the saturated steam can be calculated by  $W_{sup}^0/W_{sat}^0$ .

A summary of the numerical results for this calculation for various pressures is shown in Table III-1. The ratio  $W_{sup}^0/W_{sat}^0$  is a function of degrees of steam superheating and is not sensitive to the condition of the saturated steam. Figure III-1 shows the relation between  $W_{sup}^0/W_{sat}^0$  and the degree of superheating. Since  $W_{sup}^0/W_{sat}^0$  is not sensitive to the condition (pressure) of the saturated steam, one curve is shown in the figure.

### SUPERHEATING

The saturated steam, before entering the H<sub>2</sub>S removal reactor, may be superheated by an external energy source at constant pressure. The superheater can be a furnace which burns fuel oil or a heating coil which is powered by a part of electricity generated from the steam turbine power generator. Assuming the former, the energy required for superheating is calculated by

$$W_h = (h_{sup} - h_{sat}) / \eta_h$$

where  $W_h$ : Energy required for superheating steam, Btu/lb

$\eta_h$ : efficiency of superheater.

$\eta_h = 0.80$  are used in this study.

TABLE III-1. Change of Available Work in Throttling Process

Saturated Steam			Superheated Steam After Throttling		Change of Available Work - $\Delta W$ , Btu/lb	Ratio $\frac{W_{sup}^0}{W_{sat}^0}$	
Press. psia	Temp. °F	Available Work $W_{sat}^0$ , Btu/lb	$\Delta T$ °F	Press. psia			Available Work $W_{sup}^0$ , Btu/lb
150	358.43	283.44	5	122	270.26	13.18	0.95
			10	101	259.12	24.32	0.91
			20	78	243.30	40.14	0.85
			30	62	228.65	54.79	0.81
140	353.04	279.00	5	116	267.11	11.89	0.96
			10	98	257.11	21.89	0.92
			20	75	240.74	38.26	0.86
			30	59	226.09	52.91	0.81
130	347.33	274.13	5	109	263.47	10.66	0.96
			10	93	253.51	20.62	0.92
			20	70	236.51	37.62	0.86
			30	56	221.86	52.27	0.81
120	341.27	268.96	5	100	257.97	10.99	0.96
			10	85	248.00	20.96	0.92
			20	66	231.60	37.36	0.86
			30	53	217.53	51.43	0.81
110	334.79	263.30	5	92	252.85	10.45	0.96
			10	79	243.47	19.83	0.92
			20	62	227.07	26.23	0.86
			30	49	213.00	50.30	0.81
100	327.82	257.09	5	85	247.05	10.64	0.96
			10	74	238.26	18.83	0.93
			20	56	221.26	35.83	0.86
			30	44	206.61	50.48	0.80
90	320.28	250.15	5	77	240.46	9.69	0.96
			10	67	231.67	18.48	0.93
			20	51	215.26	34.89	0.86
			30	41	201.20	48.95	0.80
80	312.04	242.38	5	68	232.89	9.49	0.96
			10	60	224.68	17.70	0.93
			20	46	208.27	34.11	0.86
			30	37	193.62	48.76	0.80

The net available work from the steam is calculated by:

$$W_{\text{sup}}^0 = H_{\text{sup}} - h_0 - T_0 (S_{\text{sup}} - S_0) - (h_{\text{sup}} - h_{\text{sat}}) / h$$

The loss of available work  $W^0$  and the ratio  $W_{\text{sup}}/W_{\text{sat}}$  were calculated and are listed in Table III-2 for the saturated steam at the same conditions as in the throttling process. Figure III-2 summarizes the finding. The same conclusions as in the throttling process are obtained, i.e., the higher the degrees of superheating the larger the loss of the available work and  $W_{\text{sup}}/W_{\text{sat}}$  is insensitive to the condition of the saturated steam.

From Figure III-1 and Figure III-2 it can be seen that for the same degree of superheating the process which superheats the saturated steam by an external energy source is more favorable than the throttling process with respect to a minimum loss in available work.

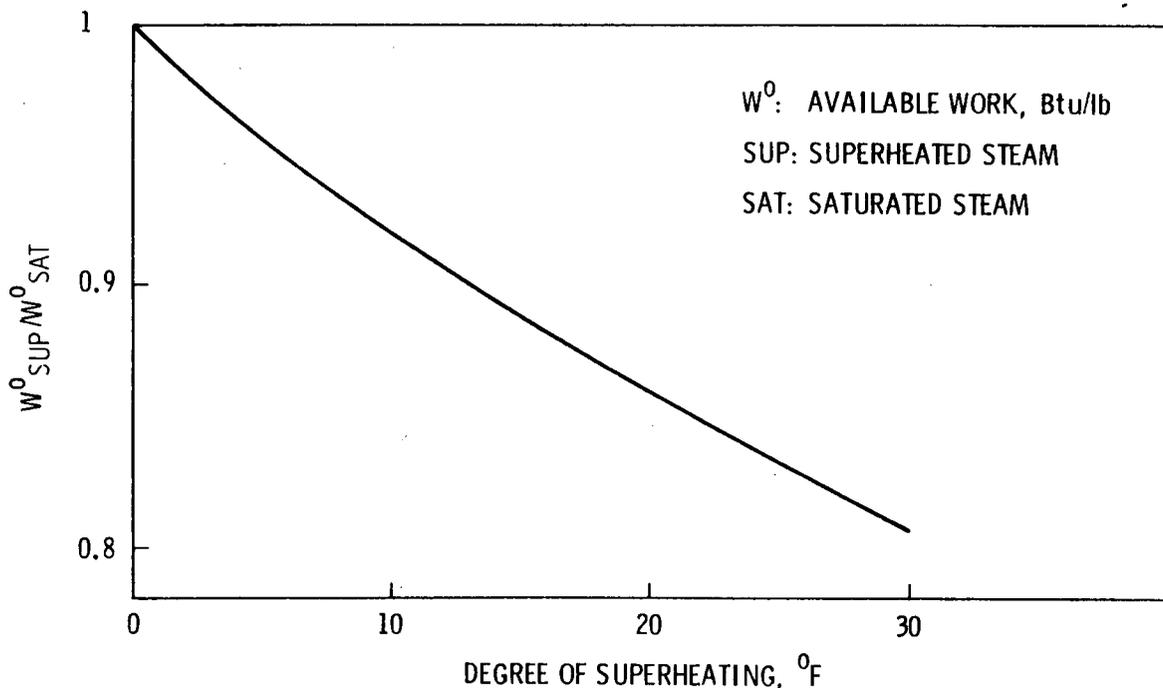


FIGURE III-1. Ratio of  $W^0_{\text{sup}}/W^0_{\text{sat}}$  vs Degrees of Superheating for the Steam Throttling Process

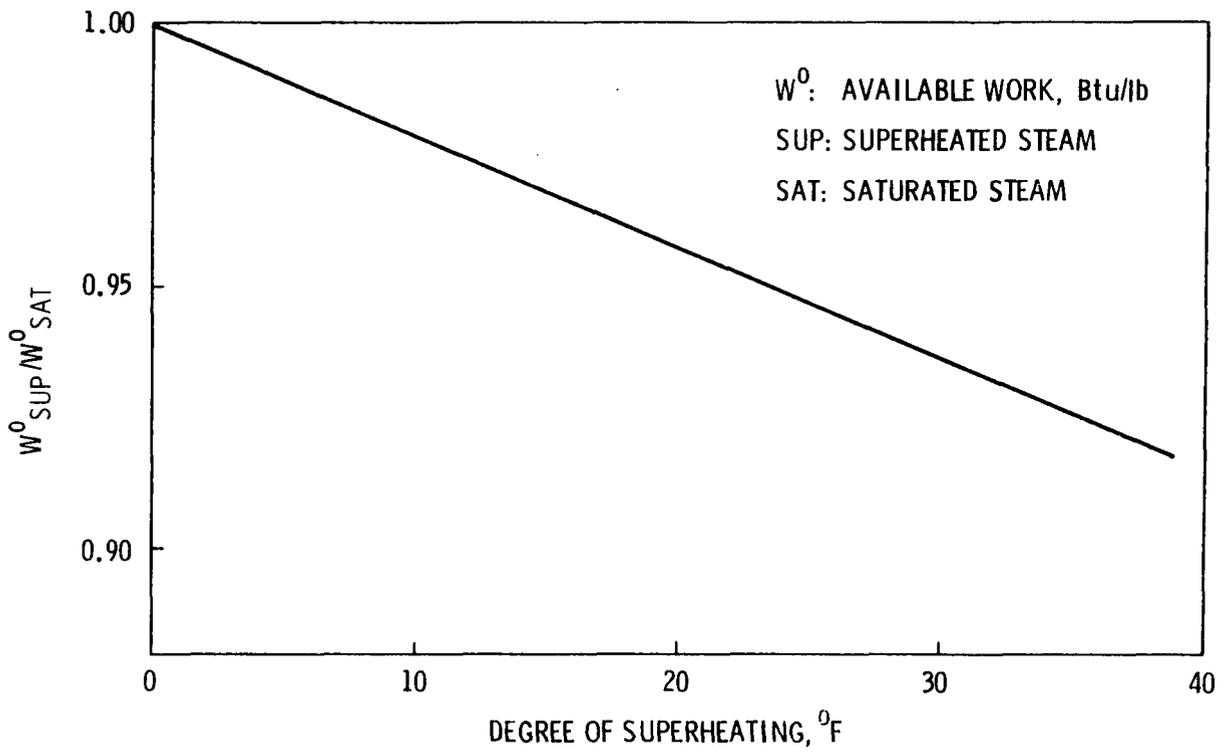


FIGURE III-2. Ratio of  $W^0_{sup}/W^0_{sat}$  vs Degree of Steam Superheating for the Steam Superheating Process

TABLE III-2. Change of Available Work in Superheating

Saturated Steam			Superheated Steam After Superheating				Change of Available Work - $\Delta W$ , Btu/lb	Ratio $\frac{W_{sup}^0}{W_{sat}^0}$
Press. psia	Temp. °F	Available Work $W_{sat}^0$ , Btu/lb	Temp. °F	$\Delta T$ °F	$W_h^{(1)}$ Btu/lb	Net Available Work $W_{sup}^0$ , Btu/lb		
150	358.43	283.44	360	1.57	1.25	282.49	0.95	0.99
			370	11.57	8.88	276.56	6.88	0.98
			380	21.57	16.50	270.82	12.68	0.96
			390	31.57	23.88	265.18	18.26	0.94
140	353.04	279.00	360	6.96	5.25	274.90	4.10	0.99
			370	16.96	12.88	269.10	9.90	0.96
			380	26.96	20.25	263.47	15.53	0.94
			390	36.96	27.50	257.91	21.09	0.92
130	347.33	274.13	350	2.67	2.13	272.52	1.61	0.99
			360	12.67	9.63	266.69	7.44	0.97
			370	22.67	17.31	260.75	13.38	0.95
			380	32.67	25.00	254.76	19.37	0.93
120	341.27	268.96	350	8.73	5.97	264.44	4.52	0.98
			360	18.73	13.75	258.24	10.72	0.96
			370	28.73	21.00	252.69	16.27	0.94
			380	38.73	28.12	247.22	21.74	0.92
110	334.79	263.30	340	5.21	3.75	260.32	2.98	0.99
			350	15.21	11.00	254.60	8.70	0.97
			360	25.21	18.25	249.04	14.26	0.95
			370	35.21	25.25	243.60	19.70	0.93
100	327.82	257.09	330	2.28	1.50	255.85	1.24	0.99
			340	12.18	8.75	250.12	6.97	0.97
			350	22.18	15.87	244.54	12.55	0.95
			360	32.18	22.87	239.16	17.93	0.93
90	320.28	250.15	330	9.72	6.88	244.66	5.49	0.98
			340	19.72	13.87	239.11	11.04	0.96
			350	29.72	20.87	233.67	16.48	0.93
			360	39.72	27.75	228.31	21.84	0.91
80	312.04	242.38	320	7.96	5.50	237.88	4.50	0.98
			330	17.96	12.50	232.38	10.00	0.96
			340	27.96	19.37	226.96	15.42	0.94
			350	37.96	26.12	231.63	20.75	0.91

APPENDIX IV

DETAILS OF COST ESTIMATION

## Details of Cost Estimation

### I. Capital Investment

#### A. Reactors

21 reactors with following specifications are required.

Diameter = 5.7 ft

Length = 3 ft

Pressure = 150 psig

Material = Stainless Steel

From "Process Plant Estimation Evaluation and Control" by Kenneth M. Guthrie, Craftsman Book Company of America, page 150, base cost of reactor = \$2,100. Shell material factor = 4.25 for solid stainless steel, pressure factor = 1.4, tray cost =  $\frac{1}{2}$  of reactor cost, and escalation index (from 1970 to 1978) = 1.85. Assume that installed cost is 150% of purchase equipment cost, then:

$$\begin{aligned}\text{Reactor Cost} &= (2100) (1.85) (4.25) (1.4) (1.5) (1.5) \\ &= \$52,010 = \$52,000/\text{reactor}\end{aligned}$$

$$\begin{aligned}\text{Total Reactor Cost} &= (52,000) (21) = \$1,092,000 \\ &\doteq \$1,100,000\end{aligned}$$

#### B. Solvent Storage Tanks

Total volume of solvent is 5,445 gal or 5,500 gal as calculated previously. Assuming that each well has two solvent storage tanks, the volume of each tank is calculated as:

$$\begin{aligned}5,500/(7) (2) &= 392.8 \text{ gal} \doteq 400 \text{ gal} \\ &= 53.5 \text{ ft}^3\end{aligned}$$

Assuming that cylindrical storage tanks with  $D = h$  are used, the diameter of tank can be calculated as:

$$V = h D^2 \frac{\pi}{4} = \frac{D^3 \pi}{4}$$

$$D = \frac{(4V)^{1/3}}{\pi} = \left[ \frac{(53.5) (4)}{\pi} \right]^{1/3} = 4.08 \text{ ft} = 4 \text{ ft}$$

From "Process Plant Estimating Evaluation and Control" page 151 the following data are obtained:

Base cost = \$1,500/tank

Material factor = 4.25 for solid stainless steel

Pressure factor = 1.0

Escalation Index = 1.85

Assuming that the installed equipment cost is 150% of purchased equipment cost, the installed equipment cost is:

$$(1,500) (4.25) (1) (1.85) (1.5) = \$17,690/\text{tank} \\ \doteq \$18,000/\text{tank}$$

Total cost of solvent storage tanks is:

$$(18,000) (14) = \$252,000$$

#### C. Air Compressor

The capacity of air compressor is:

$$(15,125) \left(\frac{1}{29}\right) (359) \left(\frac{1}{24}\right) \left(\frac{1}{60}\right) \left(\frac{520}{492}\right) = 137.2 \text{ ft}^3/\text{min} \\ \doteq 140 \text{ ft}^3/\text{min}$$

From "Plant Design and Economics for Chemical Engineers," Peters and Timmerhaus, McGraw-Hill, page 469, the purchased cost of the air compressor is \$11,500

Assuming that the escalation index from 1967 to 1978 is 2, and installed equipment cost is 170% of purchased equipment cost, the cost of air compressor is:

$$(11,500) (2) (1.7) = \$39,100 = \$40,000$$

#### D. Heat Exchangers

To calculate the size of heat exchangers the following assumptions and data are needed:

Dichloroethane:  $C_p = 0.3 \text{ Btu/lb/}^\circ\text{F}$   
 $P = 1.26$   
 $\mu = 0.9 \text{ centipoise}$   
 $k = 0.094 \text{ Btu/(hr) (ft}^2\text{) (}^\circ\text{F/ft)}$

Volume of solvent:  $(34.67) (7) = 242.69 \text{ ft}^3$ . If 5 minutes is required to heat solvent to the desired temperature, the flow rate of solvent is:  $G = (242.69) (30.48)^3 (1.26) \left(\frac{1}{453.6}\right) \left(\frac{60}{5}\right) = 229073.85 \text{ lb/hr}$

$$\begin{aligned} \text{Heat Exchanger tube ID} &= 0.584'' \\ \text{OD} &= 0.75'' \\ X_w &= 0.083'' \\ K_w &= 26 \text{ Btu}/(\text{hr}) (\text{ft}^2) (^\circ\text{F}/\text{ft}) \end{aligned}$$

$$N_{re} = \frac{DG}{\mu} = \frac{(0.584) (229073.85)}{(12) (0.9) (2.42)} = 5,118.58$$

$$\frac{C_p \mu}{k} = \frac{(0.3) (0.9) (2.42)}{0.094} = 6.95$$

$$h = \frac{k}{D} (0.023) \left(\frac{DG}{\mu}\right)^{0.8} \left(\frac{C_p \mu}{k}\right)^{1/3} \left(\frac{\mu}{\mu_w}\right)^{0.14}$$

$$= 88.09 \text{ Btu}/(\text{hr}) (\text{ft}^2) (^\circ\text{F})$$

$$h_o = \frac{k_f A}{D_o F_s} \left(\frac{D_o G_s}{\mu_f}\right)^{0.6} \left(\frac{C_p \mu}{k}\right)_f^{1/3}$$

$$= 186.16 \text{ Btu}/(\text{hr}) (\text{ft}^2) (^\circ\text{F})$$

$$\frac{1}{U_{d_i}} = \frac{1}{h_c} + \frac{D_c}{h_o D_o} + \frac{D_i X_w}{K_w D_w}$$

$$= \frac{1}{88.1} + \frac{0.584}{186 \times 0.75} + \frac{0.584 \times 0.084}{26 \times 0.667 \times 12}$$

$$= 0.0158$$

$$U_{d_i} = 63.4 \text{ Btu}/(\text{hr}) (\text{ft}^2) (^\circ\text{F})$$

Assuming that the warm solvent inlet and outlet temperatures are 80° and 30°C respectively, and that the cold solvent inlet and outlet temperatures are 15° and 65°C, respectively, the amount of heat transfer in the exchanger is:

$$q = (229073.85) (0.3) (50) (1.8) = 6184993.95 \text{ Btu/hr}$$

$$\doteq 6,185,000 \text{ Btu/hr}$$

$$A = \frac{q}{U\Delta t} = \frac{6,185,000}{(63.4) (50)} = 1951 \text{ ft}^2$$

From "Plant Design and Economics for Chemical Engineers" page 566, Figure 14-15, the purchased cost of heat exchanger is \$10,000. Installed cost is assumed as 150% of purchased cost and escalation index is 2.

Two heat exchangers, one for solvent preheating and one for solvent vapor recovery, are needed.

The total cost for heat exchangers is:

$$\text{Installed cost for heat exchangers}$$

$$= (2) (10,000) (1.5) 2 = \$60,000$$

E. Crystallizer

Estimated cost = \$40,000

F. Centrifuge

Estimated cost = \$40,000

G. Superheater

Heat required for superheating = 5.6 Btu/lb steam flow rate =  $1 \times 10^6$  lb/hr = 16,666.67 lb/min. Heat rate = (16,666.67) (5.6) = 93,333 Btu/min from "Process Plant Estimating Evaluation and Control"

Base cost = \$50,000

$$F_d = 1, F_m = 0.354, F_p = 0 \text{ and escalation index} = 1.78. \text{ Cost of}$$

$$\text{furnace} = 50,000 [1 + 0.354 + 0] (1.78)$$

$$= \$120,506$$

and the installed equipment cost is (120.506) (2) = \$241,012  $\doteq$  \$245,000.

## H. Cost of Activated Carbon

The price of activated carbon (Oct. 30, 1978) quoted from the manufacturer is shown in the following table:

Range ft <sup>3</sup>	Sample 5			
	12 x 30		4 x 10	
	\$/ft <sup>3</sup>	\$/lb	\$/ft <sup>3</sup>	\$/lb
0 to 60	46.00	1.38	51.00	1.50
60 to 299	44.25	1.30	49.00	1.44
300 to 899	43.00	1.26	47.67	1.40
More than 900	42.00	1.24	46.57	1.37

The amount of activated carbon required is 903.6 lb/reactor as calculated previously.

Based on the price of carbon given in the above table, \$1.37/lb, the total cost of catalyst is:

$$(903.6) (21) 1.37 = \$25,996 \doteq \$26,000$$

## I. Cost of Solvent

The cost of dichloroethane quoted from "Chemical Marketing Report" (Oct. 16, 1978) is \$0.11 and \$0.12/lb by Tank car, low and high, respectively. The amount of solvent required for the extraction is:

$$(34.67) (7) (3) = 728.07 \text{ ft}^3$$

If the specific gravity of solvent is 1.257 and the price of it is \$0.12/lb, the cost of solvent is:

$$(728.07) (1.257) (62.4) (0.12) = \$6852 \doteq \$7000$$

## II. Operating Costs

### A. Labor Cost:

The following manpower and labor costs are used for the estimation:

Four operators	\$19,500/year/man
One maintenance	\$21,500/year/man
One supervisor	\$23,500/year/man

$$\begin{aligned} \text{Total labor cost} &= (19,500) (4) + 21,500 + 23,500 \\ &= \$123,000/\text{year} \end{aligned}$$

B. Electricity Cost:

Work required for air compressor

$$-W = \frac{n}{n-1} P_1 V_1 \left[ \left( \frac{P_2}{P_1} \right)^{\frac{n-1}{n}} - 1 \right]$$

where  $n = 1.34$

$$P_1 = 14.7 \text{ psi}$$

$$P_2 = 150 \text{ psi}$$

$$V_1 = (359) \left( \frac{520}{492} \right) \left( \frac{1}{29} \right) = 13.08 \text{ ft}^3/\text{lb}$$

$$-W = \frac{1.34}{1.34-1} (14.7) (144) (13.08) \left[ \left( \frac{150}{14.7} \right)^{\frac{1.34-1}{1.34}} - 1 \right]$$

$$= 92430.36 \text{ ft-lb-force/lb-mass}$$

Theoretical power requirement is:

$$(92430.36) \left( \frac{15125}{24 \times 60} \right) \left( \frac{1}{33000} \right) = 29.24 \text{ hp}$$

Actual power requirement is:

$$29.24/0.85 = 34.61 \text{ hp} \doteq 35 \text{ hp}$$

Or energy requirement is:

$$35 \times 2544.9/3412.87 = 26.1 \text{ kWh}$$

If the electricity required for offsite facilities and auxiliary equipment is equal to 100% of the energy required for the air compressor and the cost of energy is 15 mil/kWh, the total cost for electricity is:

$$\begin{aligned} (26.1) (2) (365) (24) (0.75) (0.015) &= \$5,144/\text{year} \\ &= \$5,150/\text{year} \end{aligned}$$

### C. Fuel Cost

Saturate steam at 115 psig (338.08°F) is to be superheated to 345°F. From steam table the energy required for superheating is calculated as:

$$(1203.8 - 1189.8) \left(\frac{7}{21.92}\right) \left(\frac{1}{0.8}\right) = 5.59 \text{ Btu/lb} \\ \doteq 5.6 \text{ Btu/lb}$$

If the cost of heating oil is \$3.00/10<sup>6</sup> Btu, the fuel cost is:

$$(5.6) (1) (10^6) (365) (24) (0.75) \left(\frac{3}{106}\right) = \$110,376 = \text{year} \\ \doteq \$110,500/\text{year}$$

### D. Cost of Make-up Solvent

If the amount of solvent loss during the sulfur extraction is assumed as 4% of carbon treated, the solvent loss is calculated as:

$$\text{solvent loss} = (12,650) \left(\frac{1}{3}\right) (4) (0.04) \\ = 674.7 \text{ lb/day}$$

$$\text{Cost of make-up solvent} = (674.7) (0.12) (365) (6.75) \\ = \$22,164/\text{year}$$

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