RESEARCH TRIANGLE INSTITUTE

May 1998

Engineering Evaluation of Hot-Gas Desulfurization with Sulfur Recovery

Topical Report

Work performed under
Contract No. DE-AC21-94MC31258--19

for
U.S. Department of Energy
Federal Energy Technology Center
3610 Collins Ferry Road
Morgantown, WV 26505

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ABSTRACT

Engineering evaluations and economic comparisons of two hot-gas desulfurization (HGD) processes with elemental sulfur recovery, being developed by Research Triangle Institute, are presented. In the first process, known as the Direct Sulfur Recovery Process (DSRP), the SO_2 tail gas from air regeneration of zinc-based HGD sorbent is catalytically reduced to elemental sulfur with high selectivity using a small slipstream of coal gas. DSRP is a highly efficient first-generation process, promising sulfur recoveries as high as 99% in a single reaction stage. In the second process, known as the Advanced Hot Gas Process (AHGP), the zinc-based HGD sorbent is modified with iron so that the iron portion of the sorbent can be regenerated using SO_2. This is followed by air regeneration to fully regenerate the sorbent and provide the required SO_2 for iron regeneration. This second-generation process uses less coal gas than DSRP. Commercial embodiments of both processes were developed. Process simulations with mass and energy balances were conducted using ASPEN Plus. Results show that AHGP is a more complex process to operate and may require more labor cost than the DSRP. Also capital costs for the AHGP are higher than those for the DSRP.

However, annual operating costs for the AHGP appear to be considerably less than those for the DSRP with a potential break-even point between the two processes after just 2 years of operation for an integrated gasification combined cycle (IGCC) power plant using 3 to 5 wt% sulfur coal. Thus, despite its complexity, the potential savings with the AHGP encourage further development and scaleup of this advanced process.

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ACKNOWLEDGMENTS

This research was sponsored by the Federal Energy Technology Center of the U.S. Department of Energy. Valuable guidance and suggestions provided by the Contracting Officer's Representative, Mr. Thomas P. Dorchak, are sincerely acknowledged.

EXECUTIVE SUMMARY

INTRODUCTION

Hot-gas desulfurization (HGD) of coal gas in integrated gasification combined cycle (IGCC) power systems has received a great deal of attention over the past two decades due to the potential for high thermal efficiency (up to 47%) and low environmental impact of these advanced power systems. In an advanced IGCC system, coal is gasified at elevated pressures, typically 20 to 30 atm, to produce a low-volume fuel gas which is desulfurized prior to burning in a combustion turbine to produce electricity. Higher efficiency and lower cost are achieved by efficient air and steam integration, and modular designs of the gasification, hot-gas cleanup, and turbine subsystems (Figure E-1). Hot gas cleanup primarily involves removal of particulates and sulfur—mostly hydrogen sulfide (H_2S) and some carbonyl sulfide (COS). H_2S and COS can be efficiently removed to less than 20 ppmv at 350 to 650 °C using zinc-based metal oxide sorbents that can be regenerated for multicycle operation.

Air regeneration of these sorbents results in a dilute sulfur dioxide (SO_2)-containing tail gas that needs to be disposed. Options include conversion of the SO_2 to calcium sulfate using lime (or limestone) for landfilling or conversion to saleable products such as sulfuric acid or elemental sulfur. Elemental sulfur, an essential industrial commodity, is an attractive option because it is the lowest volume product and can be readily stored, disposed, transported, and/or sold.

Research Triangle Institute (RTI), with U.S. Department of Energy (DOE) sponsorship, is pursuing the development of two processes for elemental sulfur production in conjunction with

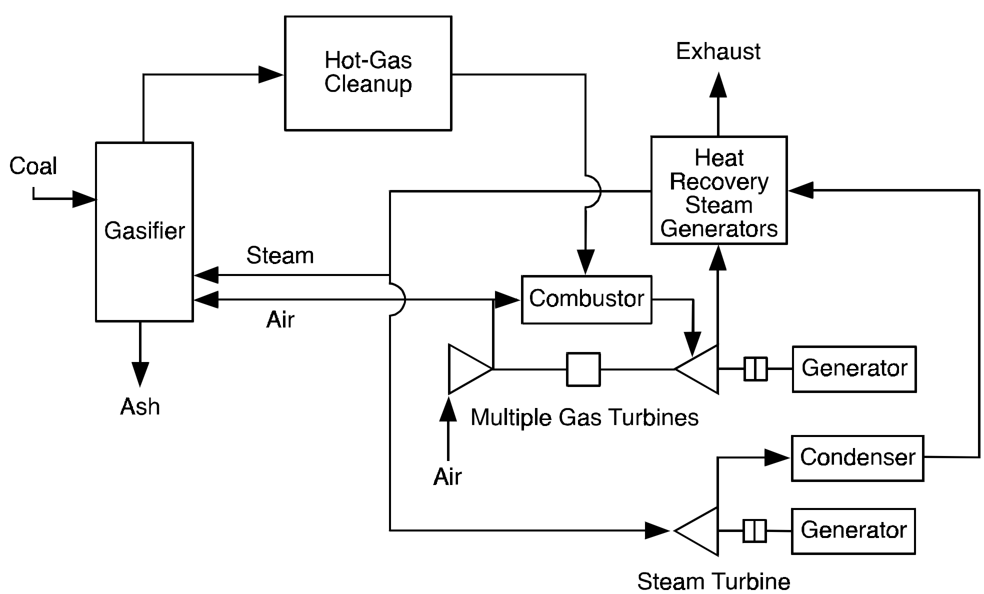

Figure E-1. Advanced IGCC system.

hot-gas desulfurization. The first process, called the Direct Sulfur Recovery Process (DSRP), involves the selective catalytic reduction of the SO_2 tail gas to sulfur using a small slipstream of the coal gas. DSRP is a highly efficient process that can recover up to 99% of SO_2 as elemental sulfur in a single catalytic reactor. However, for every mole of sulfur produced two moles of hydrogen (H_2) and/or carbon monoxide (CO) are consumed in DSRP and this represents an energy penalty for the IGCC plant. DSRP is currently in an advanced state of development.

A second-generation process being pursued by RTI involves the use of a modified zinc-based sorbent (containing zinc and iron). This sorbent can be regenerated using SO_2 and O_2 to directly produce sulfur. This process, called the Advanced Hot-Gas Process (AHGP), is expected to use much less coal gas than DSRP. DSRP is currently at the pilot-plant scale development stage, whereas AHGP has been demonstrated at small bench-scale. Both DSRP and AHGP are scheduled for slipstream testing at DOE's Power Systems Development Facility (PSDF), Wilsonville, Alabama, in 1999.

OBJECTIVE

The objective of this report is to develop process simulations with mass and heat balances for the DSRP and AHGP and to provide a **preliminary** economic comparison of the two processes in conjunction with an IGCC power plant employing HGD. The process simulation and economic evaluation were carried out by RTI's subcontractor, North Carolina State University (NCSU). NCSU's report of this work in its entirety is attached as an appendix. Background, brief process description, and important results and conclusions are provided below as a stand-alone executive summary.

BACKGROUND

Sorbent Development

Research on HGD methods for coal gas in IGCC systems has concentrated on the use of regenerable metal oxide sorbents (Gangwal, 1991, 1996; Gangwal et al., 1993, 1995; Harrison, 1995; Jalan, 1985; Thambimuthu, 1993). This research and development effort has been spearheaded by DOE's Federal Energy Technology Center (FETC) and its predecessor agencies since 1975.

The HGD process using a regenerable metal oxide (MO) sorbent is typically carried out in a two-reactor system consisting of a desulfurizer and an air regenerator

$$MO + H_2S \rightarrow MS + H_2O \qquad \text{(desulfurizer)}$$
$$MS + (3/2)\, O_2 \rightarrow MO + SO_2 \qquad \text{(regenerator).}$$

The main requirement of the metal oxide sorbent is that it should selectively react with H_2S and COS in a reducing fuel gas at desired conditions (2 to 3 Mpa, 350 to 750 °C). The thermo-dynamics of the reaction should be favorable enough to achieve the desired level of H_2S and COS removal (as much as 99% or more). The metal oxide should be stable in the reducing gas environment, i.e., reduction of MO to M should be slow or thermodynamically unfavorable since it leads to loss of valuable fuel gas and could also lead to volatile metal evaporation and decrepitation of sorbent structure.

The principle requirement during air regeneration is that the sorbent should predominantly revert back to its oxide rather than to sulfate ($MO + SO_2 + 1/2\ O_2 \rightarrow MSO_4$). Air regeneration is highly exothermic and requires tight temperature control using large quantities of diluent (N_2) or other means to prevent sorbent sintering and sulfate formation.

The bulk of research on regenerable sorbents has been on zinc-based sorbents because sorbents based on zinc oxide appear to have the fewest technical problems among all sorbents. Zinc oxide (ZnO) has highly attractive thermodynamics for H_2S adsorption and can reduce the H_2S to parts-per-million levels over a very wide temperature range. Iron oxide appears to be the most popular sorbent for use at around 400 °C.

A combined ZnO-iron oxide (Fe_2O_3) sorbent, namely, zinc ferrite ($ZnFe_2O_4$) was developed by Grindley and Steinfeld (1981) to combine the advantages of ZnO and Fe_2O_3. A temperature range of 550 to 750 °C received the major research emphasis in the United States during the 1980s and early 1990s. Because of zinc oxide's potential for reduction ($ZnO + H_2 \rightarrow Zn + H_2O$) at >600 °C followed by evaporation, a zinc oxide-titanium oxide sorbent, namely zinc titanate sorbent, was developed and tested at high temperature and high pressure (HTHP) (Gangwal et al., 1988). Zinc titanate is currently one of the leading sorbents.

During recent years, research emphasis has shifted toward lower temperatures (350 to 550 °C) based on a study in the Netherlands (NOVEM, 1991). According to this study, the thermal efficiency of an 800-MWe IGCC plant increased from 42.75% using cold-gas cleanup to 45.14% using HGD at 350 °C and to 45.46% using HGD at 600 °C. The small efficiency increase from 350 to 600 °C suggested that temperature severity of HGD could be significantly reduced without much loss of efficiency.

Reactor and Systems

A two-reactor configuration is necessary for HGD due to its cyclic nature. Early developments emphasized fixed beds. The highly exothermic regeneration led to a move away from fixed beds toward moving beds (Ayala et al., 1995; Cook et al., 1992) and fluidized beds (Gupta and Gangwal, 1992). Two DOE Clean Coal Technology IGCC demonstration plants, namely TECO and Sierra-Pacific, employing General Electric's (GE's) moving-bed HGD reactor system and M.W. Kellogg's transport reactor HGD system, respectively, are scheduled to begin operation this year. Fluidized-bed HGD systems are receiving a lot of emphasis due to several potential advantages over fixed- and moving-bed reactors, including excellent gas-solid contact, fast kinetics, pneumatic transport, ability to handle particles in gas, and ability to control the highly exothermic regeneration process. However, an attrition-resistant sorbent that can withstand stresses induced by fluidization, transport, chemical transformation, and rapid temperature swings must be developed.

Development of an iron-oxide sorbent-based fluidized-bed HGD reactor system has been carried out in Japan over the past several years (Sugitani, 1989). The process is now up to 200 tons of coal per day. The sorbent is prepared by crushing raw Australian iron oxide which is inexpensive, but attrition is a big problem with this sorbent. Durable zinc titanate and other zinc-based sorbent development is ongoing for application at the Sierra-Pacific plant for Kellogg's transport reactor (Gupta et al., 1996, 1997; Jothimurugesan et al., 1997; Khare et al., 1996).

A schematic of Kellogg's transport reactor system at Sierra-Pacific is shown in Figure E-2. This technology represents a significant development in HGD because it allows regeneration with neat air. Neat air regeneration produces a more concentrated SO_2 tail-gas stream containing around 14 vol% SO_2.

The initial sorbent tested at Sierra-Pacific was Phillips Z-Sorb III. Its attrition resistance was not acceptable. Phillips is continuing efforts to improve their sorbent. Recently RTI and Intercat have provided a much more attrition-resistant zinc titanate sorbent, EX-SO3, to Sierra-Pacific for testing after qualifying it through a series of bench- and process development unit (PDU)-scale tests (Gupta et al., 1997). This sorbent has been circulated in the system and has demonstrated satisfactory attrition resistance. Chemical reactivity tests with the sorbent are to be conducted shortly after the Sierra coal gasifier is fully commissioned and begins smooth operation.

Direct Sulfur Recovery Process

The patented DSRP being developed by RTI is a highly attractive option for recovery of sulfur from regeneration tail gas. Using a slipstream of coal gas as a reducing agent, it efficiently converts the SO_2 to elemental sulfur, an essential industrial commodity that is easily stored and transported. In the DSRP (Dorchak et al., 1991), the SO_2 tail gas is reacted with a slipstream of coal gas over a fixed bed of a selective catalyst to directly produce elemental sulfur at the HTHP conditions of the tail gas and coal gas. Overall reactions involved are shown below:

$$2\,H_2 + SO_2 \rightarrow (1/n)\,S_n + 2\,H_2O$$

$$2\,CO + SO_2 \rightarrow (1/n)\,S_n + 2\,CO_2$$

$$CO + H_2O \rightarrow H_2 + CO_2$$

$$H_2 + (1/n)\,S_n \rightarrow H_2S$$

$$2\,H_2S + SO_2 \rightarrow (3/n)\,S_n + 2\,H_2O\ .$$

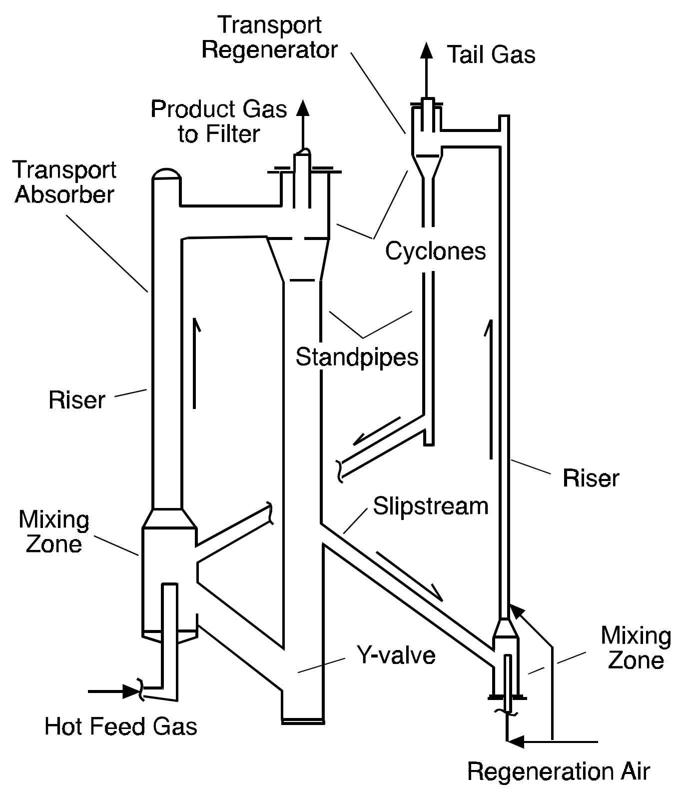

Figure E-2. Schematic of Sierra hot-gas desulfurization system.

RTI constructed and commissioned a mobile laboratory for DSRP demonstration with actual coal gas from the DOE-Morgantown coal gasifier. Slipstream testing using a 1-L fixed-bed of DSRP catalyst with actual coal gas (Portzer and Gangwal, 1995; Portzer et al., 1996) demonstrated that, with careful control of the stoichiometric ratio of the gas input, sulfur recovery of 96% to 98% can be consistently achieved in a single DSRP stage. The single-stage process, as it is proposed to be integrated with a metal oxide sorbent regenerator, is shown in Figure E-3. With the tail-gas recycle stream shown in the figure, there are no sulfur emissions from the DSRP. RTI also demonstrated the ruggedness of the DSRP catalyst by exposing it to coal gas for over 250 hours in a canister test.

The results show that, after a significant exposure time to actual coal gas, the DSRP catalyst continues to function in a highly efficient manner to convert SO_2 in a simulated regeneration tail gas to elemental sulfur. This demonstration of a rugged, single-stage catalytic process resulted in additional online experience and the assembling of more process engineering data. The development of the DSRP continues to look favorable as a feasible commercial process for the production of elemental sulfur from hot-gas desulfurizer regeneration tail gas.

Canisters of fixed-bed DSRP catalyst have been prepared for another exposure test with actual coal gas, this time at FETC's PSDF at Wilsonville, Alabama. Exposure is expected to take place sometime during FY 2000.

Additional development and testing of a fluidized-bed process is planned, capable of producing elemental sulfur from 14 vol% SO_2 at HTHP. These tests intend to demonstrate the use of DSRP in conjunction with the Kellogg transport regenerator producing 14 vol% SO_2. Due to the exothermic nature of the DSRP reactions, a fluidized-bed reactor is a preferred configuration at these high SO_2 concentrations. Two candidate attrition-resistant fluidizable DSRP catalysts have been prepared in cooperation with a catalyst manufacturer. A series of tests was conducted using these catalysts with up to 14 vol% SO_2 tail gas, at pressures from 1.0 to 2.0 Mpa, temperatures

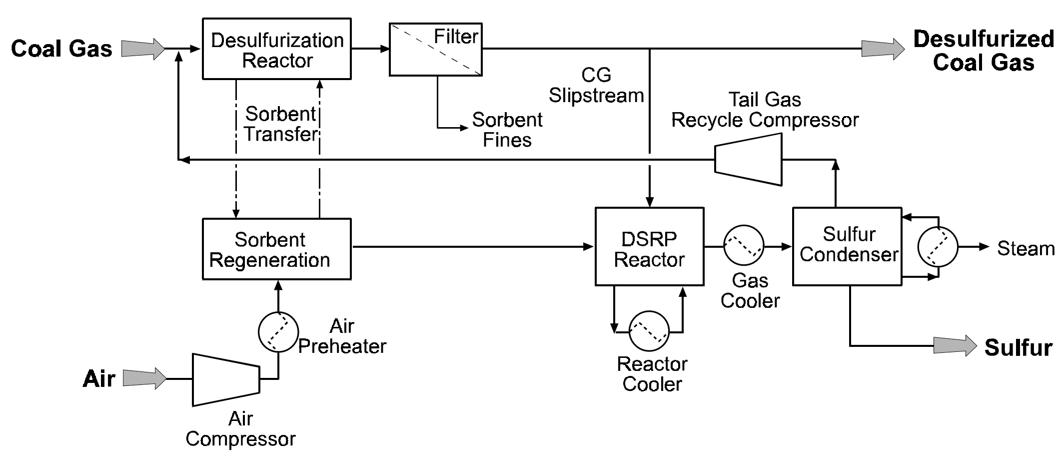

Figure E-3. Hot-gas desulfurization with DSRP.

from 500 to 600 °C, and space velocities from 3,000 to 6,000 stdcm^3/cm^3. Sulfur recoveries up to 98.5% were achieved during steady-state operation, and no attrition of the catalyst occurred in the fluidized-bed tests.

Planning is underway to conduct a long-duration field test using a skid-mounted six-fold larger (based on reactor volume) (6X) DSRP unit with a slipstream of actual coal gas at PSDF. The mobile laboratory will be refitted at RTI as a control room for the 6X unit and will be moved along with the skid-mounted 6X unit to Wilsonville, Alabama, for the testing to be conducted in FY 2000. This larger unit will utilize a fluidized-bed reactor and will be designed for production of up to 22 times more sulfur than the 7.5-cm I.D. bench-scale unit used in the previous slipstream tests.

Advanced Hot-Gas Process

In the DSRP, for every mole of SO_2, 2 mol of reducing components are used, leading to a small but noticeable consumption of coal gas. Novel regeneration processes that could lead to elemental sulfur without use of coal gas or with limited use of coal gas are being developed (Gangwal et al., 1996; Harrison et al. 1996). KEMA's hot-gas cleanup process (Meijer et al., 1996) uses a proprietary fluidized-bed sorbent which can remove H_2S to below 20 ppmv and can be regenerated using SO_2, O_2 mixtures to directly produce elemental sulfur. Along similar lines, a second-generation process, known as the Advanced Hot-Gas Process (AHGP), is being developed by RTI to regenerate the desulfurization sorbent directly to elemental sulfur with minimal consumption of coal gas. In this process (Figure E-4), a zinc-iron sorbent is used and the regeneration is carried out in two stages with SO_2 and O_2, respectively. The iron sulfide is regenerated by SO_2 in one stage to elemental sulfur. In the other stage, zinc sulfide and any remaining iron sulfide are regenerated by O_2 to provide the required SO_2. The sorbent is then returned to the desulfurizer.

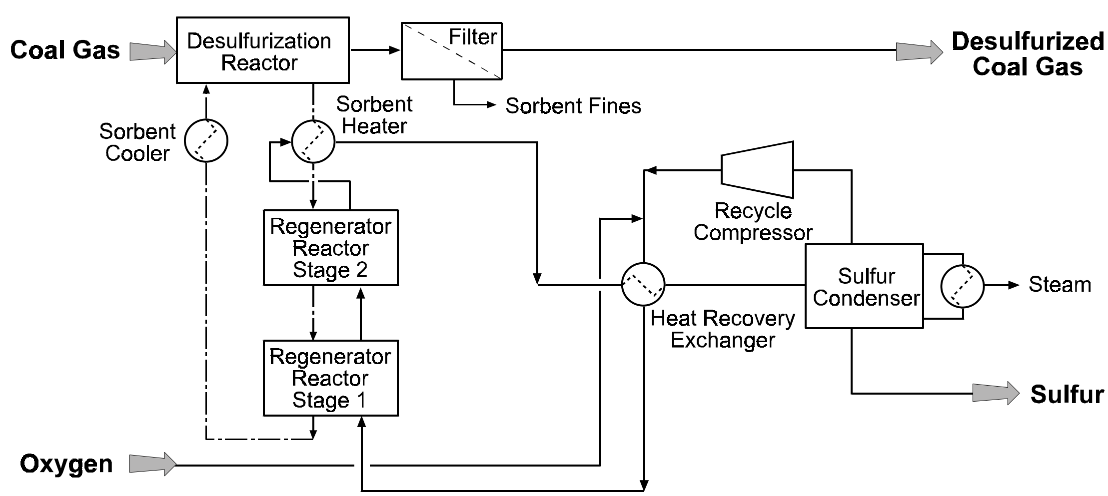

Figure E-4. Advanced hot-gas process.

The key chemical reactions of interest are as follows:

1. Sulfidation

$$Fe_2O_3 + 2H_2S + H_2 \rightarrow 2FeS + 3H_2O$$
$$ZnO + H_2S \rightarrow ZnS + H_2O$$

2. SO_2 regeneration

$$4FeS + 3SO_2 \rightarrow 2Fe_2O_3 + 7/2\ S_2$$

3. O_2 regeneration

$$2FeS + 7/2\ O_2 \rightarrow Fe_2O_3 + 2SO_2$$
$$ZnS + 3/2\ O_2 \rightarrow ZnO + SO_2.$$

The feasibility of SO_2 regeneration of combined zinc-iron sorbents was demonstrated using a thermogravimetric analyzer and high-pressure microreactor. Zinc sulfide shows essentially no SO_2 regeneration at temperatures of interest (500 to 600 °C), but zinc is needed to act as a polishing agent in the desulfurizer. A number of sorbents were prepared and tested at the bench scale over multiple cycles. Based on these tests, a highly attrition-resistant sorbent (R-5-58) was prepared and the process was demonstrated over 50 cycles in a 5.0-cm I.D. bench-scale reactor.

The results showed that R-5-58 removed H_2S down to 50 to 100 ppm levels with stable desulfurization activity over the duration. The surface area and pore volume of the sorbent did not change appreciably and the attrition index before and after the test was 3.6% and 1.2%, respectively. Sulfur balances were adequate and the SO_2 regeneration step accounted for up to 70% of the total regeneration of the sorbent. This compares to a theoretical limit of approximately 80%, assuming complete regeneration by SO_2 of the iron component.

The sorbent is being optimized further to increase its desulfurization efficiency. The goal is to develop a sorbent that can remove H_2S below 20 ppmv. Plans call for demonstrating the process at PSDF with a slipstream of actual coal gas in FY 1999 in conjunction with the DSRP field test at PDSF.

APPROACH

An engineering and economic evaluation of the DSRP (Figure E-3) and AHGP (Figure E-4) for large-scale IGCC plants was conducted using ASPEN PLUS® computer process simulation software by NCSU. The NCSU report is attached in its entirety as an appendix. Here we present a summary of the approach, key results, and conclusions.

Base case simulations of both processes assumed 0.85 mol% H_2S in the coal-gas feed. Such an H_2S concentration in the coal gas would be produced by an oxygen-blown Texaco gasification using roughly a 3.6 wt% sulfur-containing coal. Both base cases generate 260 MWe from the clean coal gas. Simulations that deviate from the base cases use suffixes to denote the changes. Table E-1 displays the significance of the suffixes. In all cases a coal-gas feed pressure and temperature of 275 psia and 482 °C, respectively, was used. However, H_2S concentration was varied from 0.25 to 2.5 mol% and power produced was varied from 110 to 540 MWe. Table E-2 shows the composition and flow rate of the raw coal gas feed to the base case HGD processes. The requirement of a higher amount of coal gas to produce the same 260 MW power by DSRP versus the AHGP is noteworthy. The DSRP was assumed to use the standard Sierra-Pacific dual transport reactor configuration shown in Figure E-2 for HGD. The DSRP reactor used for the 14% SO2 tail gas was a fast fluidized bed with an alumina-based catalyst. The AHGP reactor configuration on the other hand used a transport sulfider and a bubbling multistage fluidized-bed regenerator as shown in Figure E-5. The large bubbling reactor was required to provide a greater residence time for the slow SO_2 regeneration stage.

Table E-1. Simulation Cases Considered

Simulations	H_2S feed concentration (mol%)	MW produced
DSRP, AHGP (base cases)	0.85	260
DSRP-b, AHGP-b	2.50	260
DSRP-c, AHGP-c	0.25	260
DSRP-100, AHGP-100	0.85	110
DSRP-500, AHGP-500	0.85	540

Table E-2. Raw Gas Feed to Base Case Simulations

Component	DSRP (lb/h)	AHGP (lb/h)
H_2S	6,300	6,100
H_2O	70,500	69,000
H_2	11,800	11,500
CO	218,200	213,400
CO_2	117,400	114,800
N_2	36,300	35,500
Total	460,500	450,300

RESULTS

The preliminary process and economic evaluations conducted using ASPEN Plus are summarized. Figure E-6 compares key elements using a simple method in which each parameter for the DSRP-based process is arbitrarily assigned the value of 1.0. A range of values is produced for AHGP to cover the various cases being considered. The big advantage of the AHGP is clearly the reduced parasitic consumption of coal gas. The other operating cost elements are also lower for AHGP, because that process has a considerably lower compression power requirement. A desulfurization process based on the DSRP requires a large flow of compressed air to provide the oxygen necessary to regenerate the sulfided sorbent, and thus has a large compressor horsepower duty. By comparison, the AHGP uses oxygen only for a smaller, polishing regeneration and, by using pure oxygen, the compression duty is lowered further. The AHGP also has the SO_2 loop recycle compressor, but its duty is quite small compared to the DSRP air compressor.

[It should be noted that in the NCSU economic analysis (Appendix) the AHGP recycle compressor duty may be understated, as the calculation was based on a rough estimate for pressure drop, not a calculated value based on a piping design. By comparison, the duty for the DSRP air compressor is primarily a function of the head pressure of the system, which is well defined.]

The value of "capital cost of all equipment" for the AHGP is higher than for the DSRP-based process, as Figure E-5 shows. The higher equipment cost is primarily due to the higher cost of the AHGP reactor vessel(s). Although there are three separate reactor steps required with the DSRP-based process, the single AHGP multistage reactor vessel(s) is larger. The larger size is primarily due to the longer residence time required for the SO_2 regeneration. [It should be noted that the NCSU cost estimates (Appendix) do not include piping costs, so that the total plant capital costs will be higher than the installed equipment costs. However, since piping costs are often estimated as a direct function of the equipment cost numbers, the ratio of the installed equipment costs for the two processes shown in the figure will approximate the ratio of the total plant costs.]

Another advantage of the DSRP is that it is the easier, more understood, process to operate. This is because balancing the SO_2 production and consumption in the AHGP may be difficult.

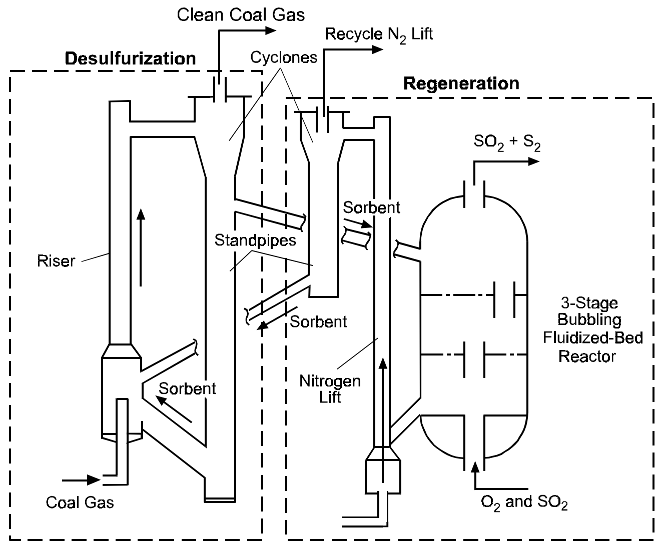

Figure E-5. Schematic of AHGP desulfurization and regeneration reactors.

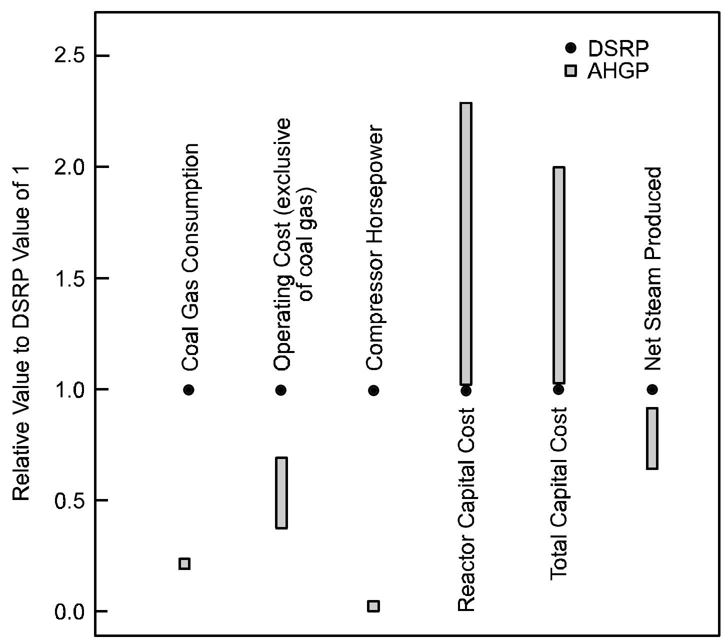

Figure E-6. Comparison of key elements of DSRP and AHGP.

Although the AHGP has a higher initial cost, indicated by its larger capital requirements, it has a significantly lower annual operating cost than DSRP. As shown in Figure E-7, the operating cost advantage of the AHGP increases as the sulfur to be recovered increases. The negative annual costs of AHGP at higher sulfur feed result from the sulfur credit with less consumption of coal gas. The operating cost difference is large enough to offset the installation cost of AHGP. As shown in Figure E-8, AHGP has a lower cumulative HGD investment after only 2 years of operation. Both Figures E-7 and E-8 are presented to illustrate only cost comparison of the two processes. Emphasis should not be placed on the accuracy of the absolute cost numbers presented in these figures.

CONCLUSIONS

ASPEN simulations of DSRP and AHGP revealed the complexity of both HGD processes. The AHGP appears to be the more difficult process to operate and may require more employees than

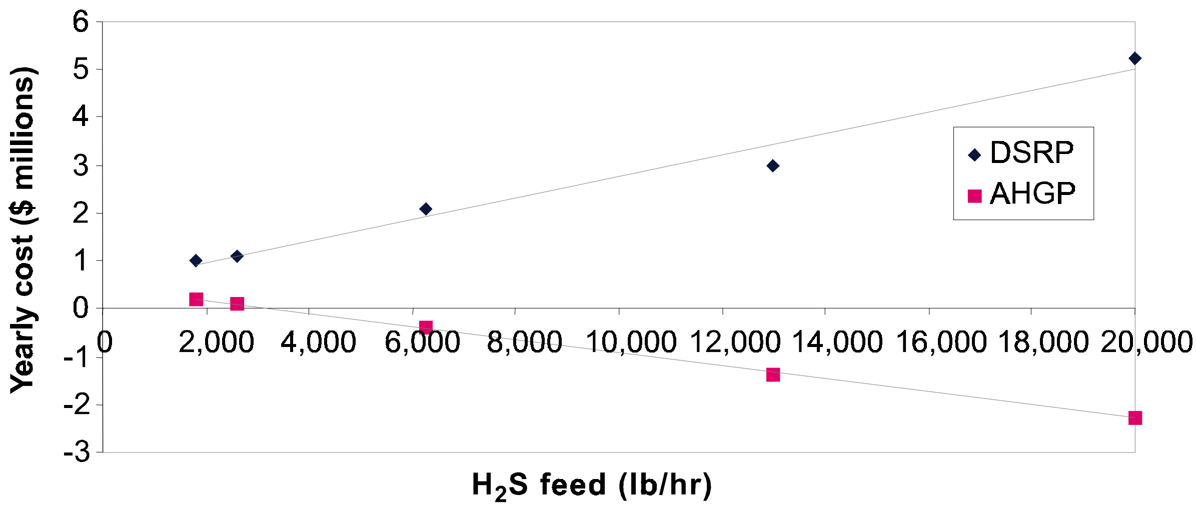

Figure E-7. Annual costs as a function of sulfur feed.

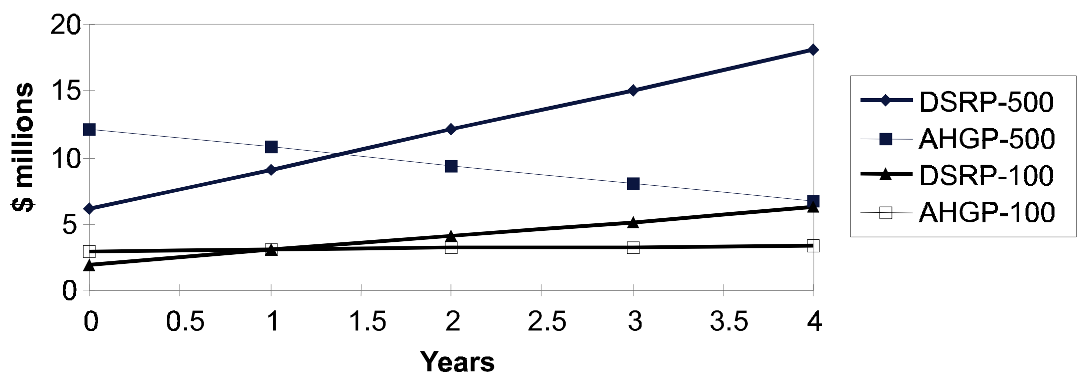

Figure E-8. Cumulative HGD investment.

the DSRP. Capital costs for the AHGP are higher than those for the DSRP—development of DSRP is also much closer to commercialization than AHGP. However, annual operating costs for the AHGP appear to be considerably less than those of the DSRP. Preliminary economic comparison shows that the total cost of implementing AHGP will be less than that of implementing DSRP after as little as 2 years of operation. Thus, despite its greater complexity, the potential savings with the AHGP encourage further development and scaleup of this advanced process.

REFERENCES

Ayala, R.E., A.S. Feitelberg, and A.H. Furman. 1995. “Development of a High-Temperature Moving-Bed Coal Gas Desulfurization System.” In *Proceedings of 12th Ann. Int. Pittsburgh Coal Conf*., p. 1053, September 11-15, Pittsburgh.

Cook, C.S., et al. 1992. “Integrated Operation of a Pressurized Fixed Bed Gasifier and Hot Gas Desulfurization System.” In *Proceedings of 12th Annual Gasif. Gas Stream Cleanup Systems Contractor’s Review Meeting*, Volume 1, DE93000228, p. 84.

Dorchak, T.P., S.K. Gangwal, and W.J. McMichael. 1991. The Direct Sulfur Recovery Process. *Environmental Progress* 19(2):68.

Gangwal, S.K. 1991. “Hot-Gas Desulfurization Sorbent Development for IGCC Systems.” IChemE Symposium Series No. 123. Sheffield, UK, pp. 159-170.

Gangwal, S.K. 1996. “Sulfur Removal from Gas Streams at High Temperature,” 3rd International Symposium on Gas Cleaning at High Temperature. University of Karlsruhe, Karlsruhe, Germany, September.

Gangwal, S.K., et al. 1988. “Bench-Scale Testing of Novel High-Temperature Desulfurization Sorbents.” Report No. DOE/MC/23126-2662 (DE89000935).

Gangwal, S.K., R. Gupta, and W.J. McMichael. 1993. “Sulfur Control Options for IGCC Systems.” In *Proceedings of 17th Biennial Low-Rank Fuels Symposium,* University of North Dakota, Energy and Environmental Research Center, St. Louis, MO, May 10-13.

Gangwal, S.K., R. Gupta, and W.J. McMichael. 1995. “Hot-Gas Cleanup-Sulfur Recovery-Technical, Environmental, and Economic Issues,” *Heat Recovery Systems and CHP*. Vol. 15, No. 2, p. 205-214, Elsevier Science Limited.

Grindley, T., and G. Steinfeld. 1981. “Development and Testing of Regenerable Hot Coal-Gas Desulfurization Sorbents.” DOE/MC/16545-1125.

Gupta, R., and S.K. Gangwal. 1992. “Enhanced Durability of Desulfurization Sorbents for Fluidized Bed Applications—Development and Testing of Zinc Titanate Sorbents.” DOE/MC/25006-3271.

Gupta, R., B.S. Turk, and S.K. Gangwal. 1996. "Bench-Scale Development of Fluid-Bed Spray Dried Sorbents." In *Proceedings of Advanced Coal-Fired Power Systems '96 Review Meeting*, Morgantown Energy Technology Center, Morgantown, WV, July.

Gupta, R., B.S. Turk, and Albert A. Vierheilig. 1997. "Desulfurization Sorbents for Transport-Bed Applications." In *Proceedings of 1997 FETC Power Systems and Environmental Control Contractor's Meeting*, Pittsburgh, PA, July.

Harrison, D.P. 1995. "Control of Gaseous Contaminants in IGCC Processes, An Overview," In *Proceedings of 12th Ann. Int. Pittsburgh Coal Conference*, p. 1047, September 11-15, Pittsburgh.

Harrison, D.P., F.R. Groves, J.D. White, W. Huang, and A. Lopez-Oritz. 1996. "Advanced Sulfur Control Processing." In *Proceedings of Advanced Coal-Fired Power Systems '96 Review Meeting*, Morgantown Energy Technology Center, Morgantown, WV, July.

Jalan, V. 1985. "High-Temperature Desulfurization of Coal Gases." In *Acid and Sour Gas Treating Processes*, Gulf Publishing Co., Houston, TX, Nov. 7.

Jothimurugesan, K., S.K. Gangwal, R. Gupta, and B.S. Turk. 1997. "Advanced Hot-Gas Desulfurization Sorbents." In *Proceedings of 1997 FETC Power Systems and Environmental Control Contractor's Meeting*, Pittsburgh, PA, July.

Khare, G.P., G.A. Delzer, G.J. Greenwood, and D.H. Kunbicek. 1996. "Phillips Sorbent Development for Tampa Electric and Sierra Pacific." In *Proceedings of Advanced Coal-Fired Power Systems '96 Review Meeting*, Morgantown Energy Technology Center, Morgantown, WV, July.

Meijer, R., F.J.J.G. Janssen, G.L. Faring, and J.W. H. Hellendoorn. 1996. "KEMA's Hot Gas Cleanup Process." *In Proceedings of 3rd International Symposium on Gas Cleaning at High Temperature*. University of Karlsruhe, Karlsruhe, Germany, September.

NOVEM. 1991. "System Study High Temperature Gas Cleaning at IGCC Systems." Netherlands Agency for Energy and the Environment.

Portzer, J.W., and S.K. Gangwal. 1995. "Slipstream Testing of Hot Gas Desulfurization with Sulfur Recovery." In *Proceedings of the Advanced Coal-Fired Power Systems '95 Review Meeting*, pp. 220-228. DOE/METC-95/1018, Vol. 1, NTIS/DE 95009732. Springfield, VA: National Technical Information Service.

Portzer, J.W., B.S. Turk, and S.K. Gangwal. 1996. "Durability Testing of the Direct Sulfur Recovery Process." In *Proceedings of the Advanced Coal-Fired Power Systems Review Meeting July 16 B18, 1996*. (CD-ROM). U.S. Department of Energy. Morgantown, WV.

Sugitani, T. 1989. Development of Hot-Gas Desulfurization Process. *Journal of the Fuel Society of Japan* 68(9):787.

Thambimuthu, K.V. 1993. Gas Cleaning for Advanced Coal-Based Power Generation. Report by IEA Coal Research, IEACR/53, London, UK.

Appendix

Process Modeling of Hot-Gas Desulfurization

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EXECUTIVE SUMMARY

This report summarizes the process simulation work and economic evaluations that were done under contract to Research Triangle Institute to aid in the design of hot gas desulfurization (HGD) processes. Two processes were evaluated for the removal of sulfur (as H_2S) from coal gas at high temperatures, that produce elemental sulfur as a byproduct. Complete mass and energy balances were accomplished for the Direct Sulfur Recovery Process (DSRP) -based process, for various feed conditions. The Advanced Hot Gas Desulfurization Process (AHGP) was also simulated for various feed conditions. ASPEN PLUS 9.3-1 was used for simulating the processes. The mass and energy balances were used in determining the equipment requirements. Equipment requirements were used for the estimation of capital costs and yearly operating costs.

The technical feasibility of the two processes was briefly evaluated. Operating the DSRP is less complicated than operating the AHGP. The AHGP contains a SO_2 loop that is balanced by reactions that consume and generate SO_2. The reaction that consumes SO_2 is equilibrium limited, and its equilibrium fractional conversion varies substantially over the range of possible reactor temperatures.

The economic evaluation shows that the AHGP has higher capital costs than the DSRP. However, the savings the AHGP provides with lower operating costs makes it the more attractive process. The economics in this report use two key assumptions: that there is a market credit for recovered elemental sulfur, and that the coal gas consumed by the HGD has an operating cost equal to the cost of the electricity that could have been generated from it. Using these and other assumptions, the analysis shows that, after only two years the AHGP should make up for its higher capital cost. After four years, AHGP could save millions over the DSRP (savings depend on plant size and the coal's sulfur concentration).

I. INTRODUCTION

1. Background

Integrated gasification combined cycle (IGCC) power plants gasify coal and then combust the coal gas to generate power. All new power plants are required to meet federal SO_X emission limitations, currently limited to 1.2 lbs per million BTU (Jaffee). Hot-gas desulfurization (HGD) removes sulfur from coal gas before combustion. HGD has the potential of reducing the cost of electricity (COE) in IGCC plants, compared to conventional liquid absorption desulfurization.

IGCC plants gasify coal using steam and either air or oxygen. The coal gas is then combusted and passes through a gas turbine, generating power. The hot exhaust gas from the turbine is then used to generate steam, which is used for additional power generation. Coal gas is produced at high temperatures and high pressures (HTHP), typically 450 to 800^{o}C and 145 to 580 psia (Gangwal). HGD reduces the coal gas sulfur content before combustion while maintaining the coal gas at HTHP conditions. Currently, IGCC plants remove sulfur with liquid phase scrubbing. The scrubbing process cools the coal gas stream below 150^{o}C. The temperature drop reduces thermal efficiency and limits the potential electricity cost reduction that is theoretically possible with IGCC power plants. IGCC power plants using liquid phase scrubbing have COE's equivalent to those of pulverized coal-based power plants (Gangwal). HGD would give IGCC power plants a competitive advantage. Implementing HGD will increase thermal efficiency, reduce the COE, and ensure SO_2 emissions are acceptable.

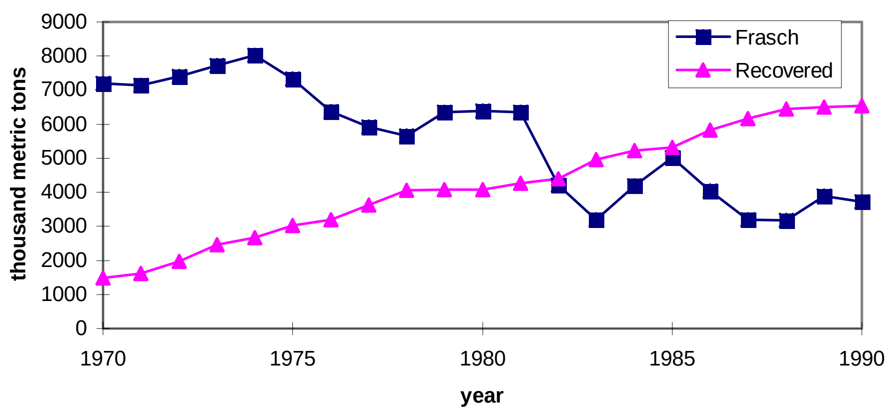
Another benefit of HGD is that the sulfur removed from the coal gas would be recovered as elemental sulfur, a valuable byproduct and easily stored material. This report describes work subcontracted to North Carolina State University (NCSU) from Research Triangle Institute (RTI). Two HGD processes that produce elemental sulfur were simulated using ASPEN PLUS 9.3-1. This work contributes to RTI efforts towards developing HGD technology. RTI research and development work includes sorbents development, characterization and a pilot-scale desulfurization testing.

Coal gas HGD and sulfur recovery could also be implemented in non-power producing applications. Although not the focus of this report, coal gas is used in methanation and Fischer-Tropsh synthesis. Methanation and Fisher-Tropsh catalysts require H_2S concentrations below 1 ppm (Cusumano) because H_2S and SO_2 poison catalysts with the formation of elemental sulfur.

2. Sulfur Production

The main purpose of the two desulfurization processes investigated is to remove sulfur from the coal gas prior to combustion, thereby reducing stack emissions. An advantage of these two processes is that elemental sulfur, which has commercial value, will be generated. Such "recovered sulfur" has been steadily replacing Frasch sulfur as a sulfur source (Figure 1). Frasch sulfur is obtained by drilling into sulfur deposits and injecting hot water, pushing molten sulfur to the surface.

Figure 1: U.S. Sulfur Production



-Data from U.S. Geological Survey

Sulfur is used in both industrial and agricultural applications. In the U.S., the majority of sulfur is used for agricultural purposes (U.S. Geological).

Recovered sulfur can be sold for $50 to $150/ton (Caruanan). Since sulfur purification was not modeled, a $50/ton credit was assigned to the recovered sulfur for the economic evaluation.

II. BASIC PROCESS DESCRIPTIONS

Two distinct desulfurization processes where simulated, the Direct Sulfur Recovery Process (DSRP) -based process and the Advanced Hot-Gas Process (AHGP). A complete collection of process flowsheets and stream summaries is contained in Appendix E. The defining characteristic of the DSRP -based process is that a slipstream of clean coal gas is used to produce the elemental sulfur from an intermediate regeneration off-gas stream containing sulfur dioxide (SO_2). The defining characteristic of AHGP is that a SO_2 stream (in a recycle loop) is used to regenerate the sorbent and produce elemental sulfur. Base case simulations for both HGD processes, referred to as "DSRP" and "AHGP", have 0.85 mol% H_2S in the coal gas feed. Both base cases also generate 260 MW from the clean coal gas. Simulations that deviate from the base cases use suffixes to denote the changes. Table 1 displays the significance of the suffixes. In all cases the coal gas feed pressure is 275 psia and its temperature is 482^oC. Simulations changes were strongly dependent on the quantity of sulfur removed from the coal gas. There is little distinction between HGD processes deviating the total sulfur removal by changing H_2S concentration and those changing sulfur removal by varying the power production.

Table 1: Coal Gas Characteristics of Simulations

Simulations	H_2S Feed Molar Concentration	MW Produced
DSRP, AHGP (base cases)	0.85 %	260
DSRP-**b**, AHGP-**b**	**2.50 %**	260
DSRP-**c**, AHGP-**c**	**0.25 %**	260
DSRP-**100**, AHGP-**100**	0.85 %	**110**
DSRP-**500**, AHGP-**500**	0.85 %	**540**

Table 2 shows the composition and flow rate of the "raw" coal gas feed to the base case HGD processes. After sulfur is removed from the streams the coal gas can produce 260 MW.

Table 2 : Raw Coal Gas Feed to Base Case Simulations

Component	DSRP (lb/hr)	AHGP (lb/hr)
H_2S	6,300	6,100
H_2O	70,500	69,000
H_2	11,800	11,500
CO	218,200	213,400
CO_2	117,400	114,800
N_2	36,300	35,500
Total	460,500	450,500

1. Direct Sulfur Recovery Process Sorbent Cycle

The term DSRP, strictly speaking , refers only to that part of the entire HGD process that produces elemental sulfur. For convenience, the process simulations were made by assuming a kind of "generic" process (Figure 2) utilizing a ZnO sorbent, with Al_2O_3 support, to remove sulfur (present in the form of H_2S) via reaction 1. The reader should note that in this report "DSRP" is often used as shorthand for the entire "DSRP-based HGD process," while the novel DSRP reactions to form elemental sulfur occur in what this report refers to as the "DSRP Reactor." Reaction 1 occurs in the desulfurization reactor (DESULF, Figure 2).

$$ZnO + H_2S \rightarrow ZnS + H_2O \qquad (1)$$

The spent sorbent is regenerated in an oxidizing environment, forming SO_2. Reaction 2 occurs in the regenerator reactor (REGEN, Figure 2), it is driven to completion by oxygen.

$$ZnS + 3/2\ O_2 \rightarrow ZnO + SO_2 \qquad (2)$$

The SO_2 exits the regenerator in a stream designated regenerator off-gas (ROG). The ROG flows to the DSRP Reactor. A slipstream of clean coal gas is also fed to the DSRP Reactor. The H_2 and CO in the coal gas slipstream participate in catalyzed reactions (3 and 4), converting SO_2

into elemental sulfur. The reactions 3 and 4 are the simplified overall reactions of a more complex series of reactions.

$$H_2 + 1/2\ SO_2 \rightarrow H_2O + 1/4\ S_2 \qquad (3)$$

$$CO + 1/2\ SO_2 \rightarrow CO_2 + 1/4\ S_2 \qquad (4)$$

The heats of reaction for converting SO_2 to elemental sulfur have been calculated by RTI (Portzer, 1996). Comparing RTI calculated values with experimental results indicated the RTI values were reasonable. Table 2 shows that ASPEN calculated heats of reaction are in general agreement with those calculated by RTI. The ASPEN model does an accurate job determining the heat evolved during reactions and therefore will predict correct heat transfer requirements in the process simulations.

Table 3: Heats of Reaction Calculated by RTI and ASPEN Model

Reaction	Temp (^oC)	ΔH_{RTI} (BTU/mole)	ΔH_{ASPEN} (BTU/mole)	difference
3	550	- 28,000	- 28,700	2.5 %
3	650	- 28,300	- 29,000	2.5 %
3	750	- 28,600	- 29,200	2.1 %
4	550	- 43,900	- 44,100	0.5 %
4	650	- 43,700	- 44,000	0.7 %
4	750	- 43,800	- 43,600	0.5 %

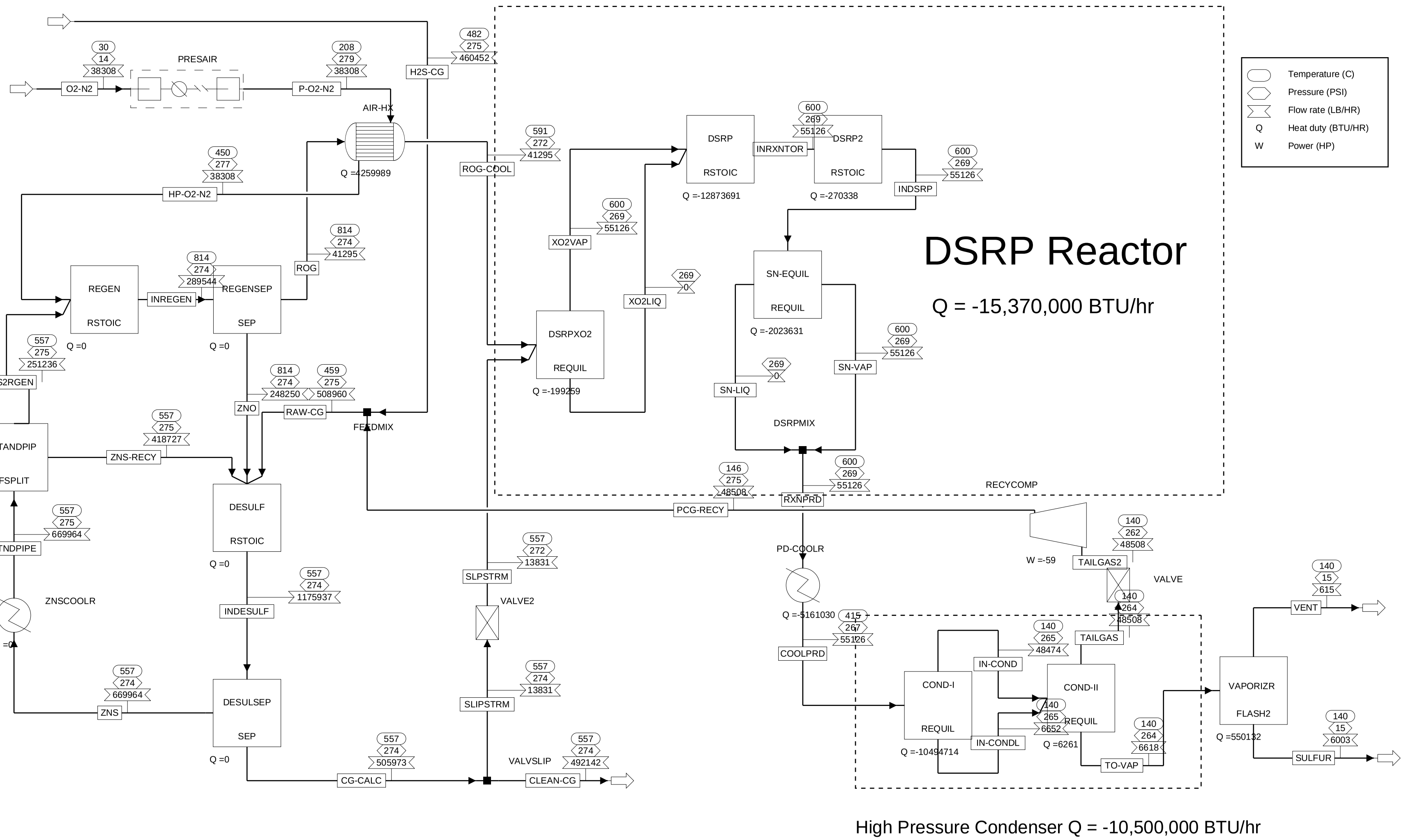
-Heat of reaction values adjusted to match stoichiometry written, P=300 psig for calculations

2. Sorbent Composition - DSRP

The oxidized sorbent, a mixture of ZnO and Al_2O_3, was assumed to contain 15 wt% zinc metal. This distribution is based on an assumed, “generic” sorbent defined by RTI, and results in an oxidized sorbent containing 18.671 wt% ZnO with the balance as inert Al_2O_3 support. While developing the process model and adjusting the stream flow rates to achieve the desired heat balance, it became desirable to increase sorbent circulation rates above the stoichiometric requirements. For these models, the ratio of Zn to Al remained unchanged. The excess Zn sorbent circulating through the system was assumed to remain in the sulfide state (ZnS).

DSRP - based Desulfurization

12/20/97 DSRP



3. Advanced Hot Gas Process Sorbent Cycle

The AHGP (Figure 3) uses a sorbent containing a mixture of ZnO and Fe_2O_3 on Al_2O_3 support for removing H_2S from the coal gas and converting it into elemental sulfur. Both zinc and iron components react with the H_2S present in the coal gas. The desulfurization reactions are represented below.

$$ZnO + H_2S \rightarrow H_2O + ZnS \quad (5)$$

$$Fe_2O_3 + 2\,H_2S + H_2 \rightarrow 2\,FeS + 3\,H_2O \quad (6)$$

The sulfided sorbent is sent to a three-stage regenerating reactor that reoxidizes the sorbent and generates elemental sulfur. Sorbent and a SO_2 gas stream flow counter-currently through the regenerator (Figure 3) (Figure 6). The sorbent enters the regenerator at the HX-STAGE (the third and highest elevated stage) where the sorbent is heated by the effluent gas stream. Sorbent descends to REGEN2 (the second stage) where SO_2 , present in great excess, oxidizes the majority of the FeS sorbent.

$$3\,SO_2 + 4\,FeS \rightleftharpoons 7/2\,S_2 + 2\,Fe_2O_3 \quad (7)$$

It has been assumed that two-thirds of the FeS oxidizes in REGEN2. Calculated equilibrium conversions for reaction 7 are listed in table 4. Sorbent enters the second stage of the regenerator at 512^{o}C and gas enters the second stage at 715^{o}C. Table 3 shows equilibrium conversions varies significantly over the range of temperatures possible in stage 2, a stage for which it is unclear what value represents its temperature the best. Simulated stage 2 exit temperatures were 580^{o}C, this exit temperature assumes perfectly mixed behavior in the stage 2. In reality there will likely be higher temperatures at lower elevations in the stage. The ASPEN model uses an RSTOICH block to simulate this stage so that the conversion can be arbitrarily fixed at 67%. This value was defined by RTI, based on experimental data. The information in Table 4 suggests that the assumed two-thirds conversion probably overestimates the actual conversion. In commercial practice, increasing the Fe:Zn ratio could compensate for lower than simulated reaction 7 conversions (conversion written in terms of FeS). Another aspect of this reactor stage is that the extent that FeS oxidizes by SO_2 will vary with temperature fluctuations and increase the difficulty in balancing SO_2 consumption and generation.

Table 4: Equilibrium Conversion for FeS Oxidation by SO_2

Regenerator Temperature (^oC)	Equilibrium Fractional Conversion
500	0.43
550	0.53
600	0.65
650	0.77
700	0.90

Equilibrium calculated from ASPEN REQUIL block, P = 275 psia

Sorbent oxidization approaches completion in the bottom regenerator stage (REGEN1, Figure 3). REGEN1 oxidizes the sorbent using pure oxygen (reactions 8 and 9). The oxidation generates SO_2, making up for SO_2 used in reaction 7.

$$7/2\ O_2 + 2\ FeS \rightarrow 2\ SO_2 + Fe_2O_3 \qquad (8)$$

$$3/2\ O_2 + ZnS \rightarrow SO_2 + ZnO \qquad (9)$$

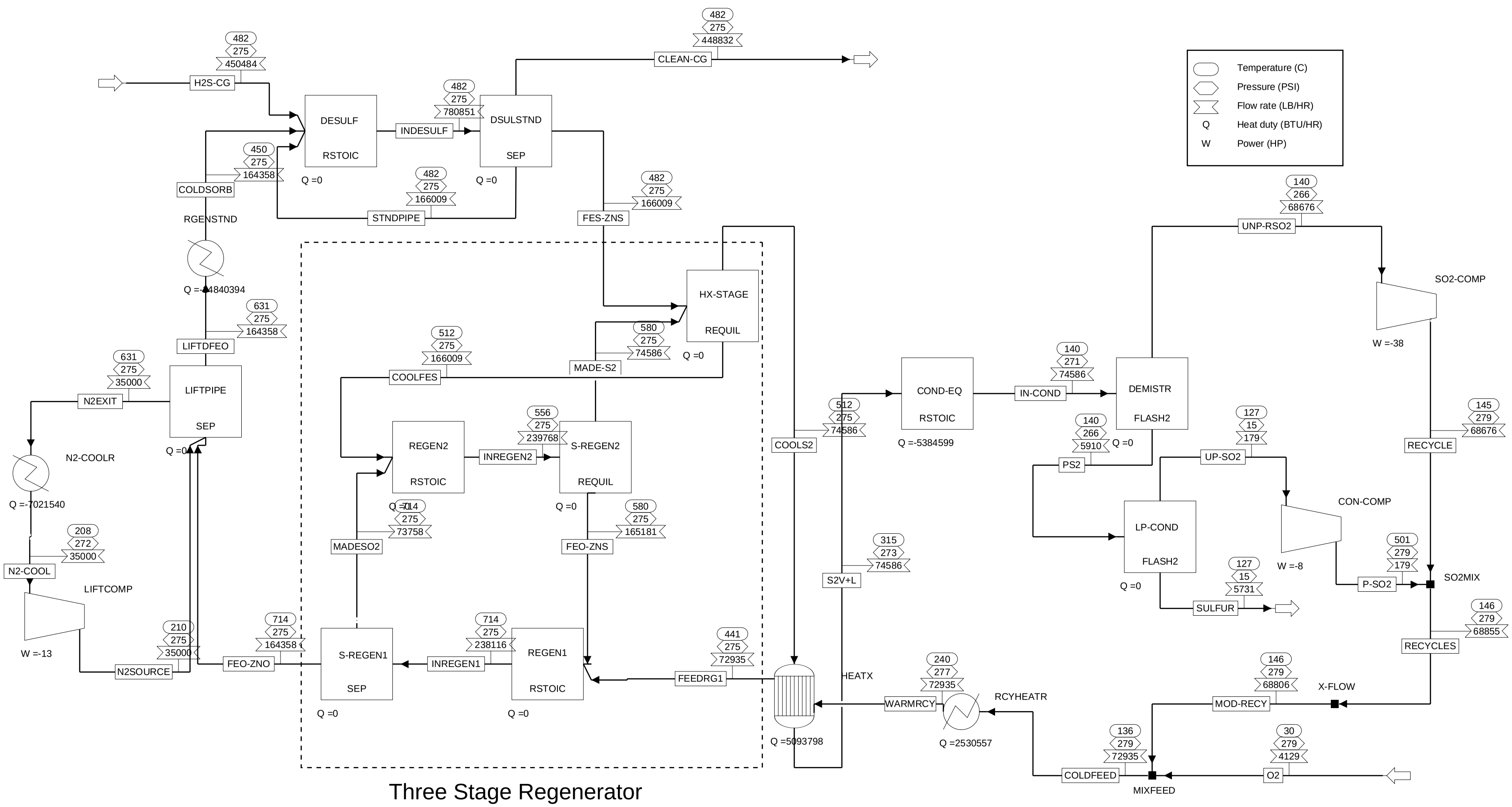
This modeling assumes that SO_2 does not oxidize sorbent in REGEN1, since equilibrium conversion for SO_2 oxidation is approached in REGEN2. The equilibrium regeneration of sorbent by SO_2 will be quickly superseded by oxygen regeneration.

4. Sorbent Composition - AHGP

AHGP sorbent composition was defined by RTI to contain 3 wt% Zn and 12 wt% Fe, which corresponds to 3.734 wt% ZnO and 17.154 wt% Fe_2O_3. The balance, 79.109 wt%, was inert Al_2O_3. As discussed above, the ratio of Fe to Zn will need to be increased if the actual conversion for reaction 7 is lower than 0.667, its assumed value.

AHGP Advanced Hot Gas Process

AHGP 1/19/98



During subsequent simulation development and adjustment of stream flow rates to achieve the desired heat balance, it became apparent the defined sorbent composition was not optimal. To run the reactors adiabatically, it was necessary to increase sorbent flow. Circulating more sorbent increased the heat capacity of the reactive stream and reduced the adiabatic temperature rise. Such a sorbent increase required an increase in Al_2O_3 flow. Increasing Fe or Zn flow would have upset the SO_2 generation and consumption balance created by reactions 7, 8 and 9. Therefore, alumina flow was increased. The effect would be the same as adding pure alumina sorbent to the reactor system, or by manufacturing a sorbent that has a lower active metal content and increasing the total flow to match the amount of alumina added.

The Al_2O_3 circulation was increased until an adiabatic regenerating reactor would operate below 716^{o}C. The effects of changing Al_2O_3 circulation ripple through the process. The required SO_2 circulation rate was affected by varying the Al_2O_3 flow. The desired SO_2 volumetric flow rate increased with increasing sorbent flow rate because of increased reactor size. Increasing the SO_2 circulation helped reduce the adiabatic temperature rise, lessening the need to increase sorbent flow. Table 5 shows how Al_2O_3 flow was increased until an acceptable adiabatic regeneration temperature was achieved. The table displays the stepwise approach used to determine the Al_2O_3 circulation needed in the AHGP-b simulation (-b signifies a 2.5 mol% H_2S in the feed). In the simulation, ZnS and FeS flow rates (leaving the desulfurization reactor) were constant at 7,600 lb/hr and 41,000 lb/hr, respectively.

Table 5: Al_2O_3 Circulation Rate Effect on Regenerator Stage 1 Temperature

Al_2O_3 (lb/hr)	T_{REGEN1} (^{o}C)	Desired SO_2 flow (ft^3/hr)
165,297	1025	102,000
330,594	787	181,000
400,000	759	214,000
450,000	715	238,000

III. PHYSICAL PROPERTIES

1. Equation of State

All simulations discussed in this report used the Peng Robinson cubic equation of state with the Boston-Mathias alpha function (PR-BM).

1.a. *Equation of State's Importance*

Modeling unit operations requires physical property information for all compounds present. In calculating thermodynamic equilibrium, fugacity coefficients are used to determine phase equilibrium. An equation of state can be used for the calculation of fugacity, as well as other important physical properties. The equation of state also relates pressure, temperature, and molar volume so that only two need to be specified and the third can be calculated. Phase equilibrium is established when the fugacity of each component is the same in all phases.
A two-phase (vapor and liquid) system is at equilibrium when:

$f^v_i = f^l_i$ i = 1,2,...N where N is the number of compounds

Where:

f^v_i	=	$\varphi^v_i y_i P$	Fugacity of component *i* in the vapor phase
f^l_i	=	$\varphi^l_i x_i P$	Fugacity of component *i* in the liquid phase

$$\ln \varphi_i^{\alpha} = -\frac{1}{RT}\int_{\infty}^{V^{\alpha}}\left[\left(\frac{\partial P}{\partial n_i}\right)_{T,V,n_{i\neq j}} - \frac{RT}{V}\right]dV - \ln Z_m^{\alpha}$$

Notation:

α	=	vapor or liquid (v or l)	P	=	Pressure
n_i	=	Mole number of component *i*	T	=	Temperature
x_i	=	Liquid mole faction of component *i*	R	=	Gas Constant
y_i	=	Vapor mole faction of component *i*	V	=	Total volume
Z	=	Compressibility factor			

The equation of state also is used to determine other properties via departure functions.

- Enthalpy departure:

$$(H_m - H_m^{ig}) = -\int_{\infty}^{V}\left(P - \frac{RT}{V}\right)dV - RT\ln\left(\frac{V}{V^{ig}}\right) + T(S_m - S_m^{ig}) + RT(Z_m - 1)$$

- Entropy departure:

$$(S_m - S_m^{ig}) = -\int_{\infty}^{V}\left[\left(\frac{\partial P}{\partial T}\right)_V - \frac{R}{V}\right]dV + R\ln\left(\frac{V}{V^{ig}}\right)$$

- Gibbs Free Energy departure:

$$(G_m - G_m^{ig}) = -\int_{\infty}^{V}\left(P - \frac{RT}{V}\right)dV - RT\ln\left(\frac{V}{V^{ig}}\right) + RT(Z_m - 1)$$

Notation:

H = Enthalpy S = Entropy G = Gibbs Free Energy

ig (superscript) denotes variable's value for ideal gas

m (subscript) denotes variable's value for the mixture

1.b. *Selection*

The Peng Robinson cubic equation of state with the Boston-Mathias alpha function (PR-BM) was used in these simulations because it was recommended for gas-processing, refinery, and petrochemical applications (ASPEN PLUS- Reference Manual 2). It was recommend for modeling nonpolar and mildly polar mixtures, including hydrocarbons and light gases like: carbon dioxide, hydrogen sulfide, and hydrogen. Reasonable results can be expected for all temperatures and pressures. The Peng-Robinson equation of state is:

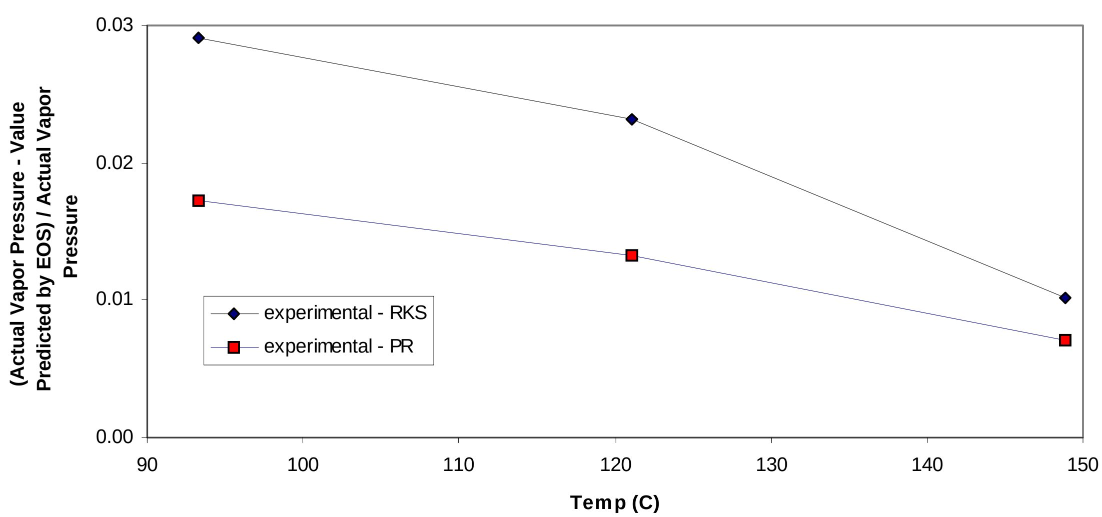
$$P = \frac{RT}{V_m - b} - \frac{a}{V_m(V_m + b) + b(V_m - b)}$$

Variables ‘a’ and ‘b’ account for attractive forces and the space occupied by all species present, R is the ideal gas constant, T is temperature and V_m is the mixture’s specific molar volume.

The Boston Mathias extrapolation is used for supercritical components. Boston and Mathias derived an alpha function that is particularly good at modeling decreasing attraction between molecules at high temperatures (ASPEN PLUS- Reference Manual 2).

The above descriptions also apply to the Redlich-Kwong-Soave cubic equation of state with Boston-Mathias alpha function (RKS-BM). The decision to use the PR-BM over RKS-BM was made after comparing literature phase data (Braker) with simulations using both property option sets. Figure 4 shows the fractional deviation of simulated vapor pressures compared to literature values. Both equations of state calculate values in good agreement with actual values, and the Peng-Robinson equation of state gives the best results.

Figure 4: RKS and PR Calculated SO2 Vapor Pressure Deviations From Tabulated Values



2. Elemental Sulfur

Accurately predicting elemental sulfur properties requires knowing which allotropes of sulfur will be formed. For the conditions occurring in the HGD process S_8, S_6, and S_2 are the predominant allotropes (Barnett; Cotton). Temperature is the dominant variable affecting the equilibrium sulfur distribution. The ASPEN simulations concurred with literature distributions, predicting S_2 predominance at high temperatures (reactor temperatures), and a shift towards S_8 and S_6 at lower temperatures (condenser temperatures). Accurate sulfur distributions are important for the integrity of phase equilibrium predictions. In addition, correctly simulating sulfur equilibrium increases the accuracy of energy balances.

It is worth noting some unusual properties of liquid elemental sulfur. Recovered sulfur should not be raised to temperatures above 159^{o}C, as above that temperature the liquid sulfur becomes increasingly viscous (Cotton). Sulfur melts around 114^{o}C; it does not have a sharp melting point due to the presence of various allotropes (Barnett).

IV. EQUIPMENT

1. DSRP- Based Process Equipment

For the purposes of this process simulation and economic evaluation, the DSRP - based HGD process was defined to have a desulfurization and regeneration transport reactor network as shown in Figure 5. Sulfur is removed from coal gas (Reaction 1) in the desulfurization reactor and sorbent regeneration (Reaction 2) takes place in the regeneration reactor. There is also a DSRP Reactor in which the elemental sulfur is formed via Reactions 3 and 4. Other major pieces of equipment in the DSRP include compressors, condensers, and heat exchangers.

$$ZnO + H_2S \rightarrow ZnS + H_2O \qquad (1)$$

$$ZnS + 3/2\ O_2 \rightarrow ZnO + SO_2 \qquad (2)$$

$$H_2 + 1/2\ SO_2 \rightarrow H_2O + 1/4\ S_2 \qquad (3)$$

$$CO + 1/2\ SO_2 \rightarrow CO_2 + 1/4\ S_2 \qquad (4)$$

In addition to Reactions 3 and 4, intermediate and side reactions occur in the DSRP Reactor. They are discussed later in the report.

1.a. *Desulfurization and Regeneration Transport Reactors - DSRP*

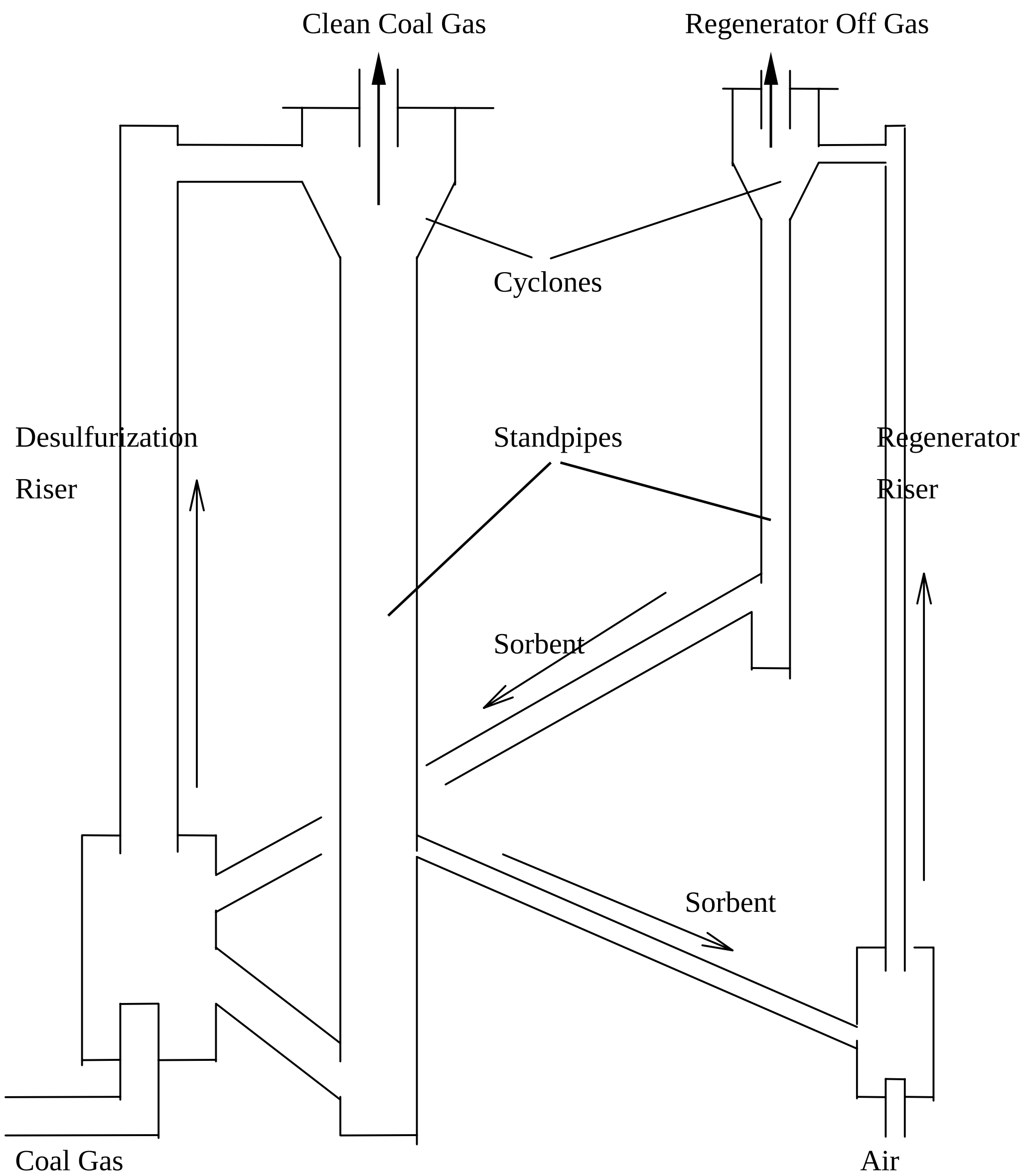
The DSRP - based HGD process is assumed to use transport reactors for the desulfurization and regeneration reactions. The Sierra Pacific hot-gas desulfurization system (Cambell) has been the basis for the reactor system design (Figure 5). Cyclones separate the sorbent from the exiting gas streams. Sorbent settles from the cyclones into standpipes. The sorbent has a relatively high residence time in the standpipes. Standpipe residence times are several minutes while reactor residence times are only several seconds long. Standpipe heat exchangers remove heat from the reactor system. During startup, sending steam through the standpipe heat exchanger could heat the sorbent partially up to reactor temperatures.

The regeneration reaction releases a substantial amount of heat. Feeding a stoichiometric amount of sorbent in the ASPEN simulation to an adiabatic regeneration reactor results in predicted temperatures surpassing 1,000^{o}C (DSRP base case). RTI guidelines stated that HGD sorbents would experience substantial sintering at temperatures above 815^{o}C. The strategy adopted to control reactor temperature is recycling excess sorbent. The additional sorbent increases the total heat capacity of the reactive streams. The additional sorbent will not result in additional reactions and the increased heat capacity will decrease the adiabatic temperature rise. The adiabatic temperature rise can be expressed by the following relationship:

$$\Delta T_{adiabatic} \approx \frac{\Delta H_{rxn}}{Cp_{stream}}$$

Increased sorbent flow was selected as the preferred strategy over that of using a reactor heat exchanger, since it simplifies reactor design. Furthermore, hot spots are more likely to occur in a reactor containing a heat exchanger. Limiting reactor temperature by reducing reactor feed stream temperatures (without additional sorbent circulation) was also investigated. This approach was discarded because the reactions would be extinguished at feed temperatures low enough to keep the reactor temperature below 815^{o}C.

Figure 5: Schematic of DSRP - Based HGD Process Desulfurization and Regeneration Reactors



The transport reactors exhibit numerous advantages over fixed-bed, fluid-bed and moving-bed reactors. The transport reactor has lower capital cost, its high flowrate of sorbent controls reactor temperatures, and the high velocities prevent hot spots from occurring on the sorbent (Campbell). The transport reactor's superior temperature control allows undiluted air to be used during regeneration.

The equations used for sizing and costing the DSRP - based process desulfurization and regeneration transport reactor system are described in Appendix G-Calculation of Reactor Size. The actual calculations can be found in Appendix H-Sizing Reactors for the DSRP.

1.b. *DSRP Reactor - DSRP*

The DSRP Reactor itself is a fast fluidized bed reactor with its catalyst modeled as Al_2O_3. There are several ASPEN blocks used to model what will be only one DSRP Reactor, a dashed box has been drawn around the series of blocks used (Figure 2). The catalyst is circulated through the reactor and an external heat exchanger. Heat is removed by cooling the catalyst while it is outside the reactor. The heat exchanger cools the catalyst to 500^oC and the catalyst is then reintroduced to the reactor at a rate that is high enough to keep the DSRP Reactor effluent near 600^oC. (Appendix D- Calculation of DSRP Catalyst Cycling Rate)

Figure 2 shows that several blocks were used for the simulation of the DSRP Reactor: DSRPXO2, DSRP, DSRP2, and SN-EQUIL.

In DSRPXO2, any oxygen that enters the DSRP as a contaminant in the ROG consumes coal gas by a conventional combustion reaction. The oxygen combines with CO forming CO_2. It is not necessary to model combustion of H_2 since the ratio of CO to H_2 will be set by the Water Gas Shift (WGS) reaction. Also in DSRPXO2 the WGS reaches equilibrium. The WGS reaction is known to reach equilibrium before the reactions of SO_2 with H_2 or CO begin (Chen, 1994). DSRPXO2 uses a Gibbs Free Energy calculation to establish equilibrium for reactions 10 and 11.

DSRPXO2	$CO + 1/2\ O_2 \rightarrow CO_2$	$X_{O2} = 1$	(10)
	$H_2O + CO = CO_2 + H_2$	$K_C(600^oC) = 2.6$	(11)

The key DSRP reactions have been modeled in the following blocks.

DSRP	$2\ H_2 + SO_2 \rightarrow 0.5\ S_2 + 2\ H_2O$	$X_{H2} = 0.99$	(3)
	$3\ CO + SO_2 \rightarrow COS + 2\ CO_2$	$X_{CO} = 0.9995$	(12)
	$H_2 + 0.5\ S_2 \rightarrow H_2S$	$X_{H2} = 0.01$	(13)
DSRP2	$SO_2 + 2\ COS \rightarrow 1.5\ S_2 + 2\ CO_2$	$X_{COS} = 0.9999$	(14)

SN-EQUIL establishes the allotropic distribution of elemental sulfur using a Gibbs Free Energy calculation. Including this block more accurately models the heat generated inside the DSRP Reactor.

SN-EQUIL	$4\ S_2 \rightarrow S_8$	$X_{S2} = 0.23$	(15)
	$3\ S_2 \rightarrow S_6$	$X_{S2} = 0.32$	(16)

1.c. *PRESAIR - DSRP*

The transport reactor design for the regenerator in the DSRP - base HGD process model allows the use of undiluted air (“neat air”) to regenerate the desulfurization sorbent. Introducing air at the required pressure can be accomplished using either an axial-flow or centrifugal compressor. In most applications, including this process simulation, it is preferable to use a centrifugal compressor. Centrifugal compressors have the advantage of a larger operating range (Dimoplom). Centrifugal compressors typically operate below 225^{o}C (Brown; Dimoplon) in order to avoid equipment damage.

The large increase in pressure (ambient to 275+ psia) in the PRESAIR air compressor generates a considerable temperature rise. Interstage cooling, between the compressor's 6 stages, is necessary to maintain an air temperature below 225^{o}C and to prevent mechanical damage to the compressor (Brown; Dimoplon). The temperature increase across the first stage does not require cooling stage 1 effluent and there is no need to cool the effluent of the final stage as well. Therefore, there will be four interstage coolers needed for the six-stage compressor. Pressure drop during interstage cooling can be approximated as 2% of the pressure entering the cooler or 2 psia, whichever is larger (Brown). For pressuring to 280 psia estimating a 2 psia drop for each cooler is reasonable; these pressure losses are included in the ASPEN PLUS compressor block calculations.

Significant capital will be spent on the purchase of an air compressor. Increasing pressure to 280 psia for an feed of 8,800 ft^3/min (DSRP base case) requires a compressor made of steel as opposed to cast iron (Bloch). Compressors made of low value steel should be both mechanically durable and economical. For simplicity, the cost estimates in this report assume electric drive.

Steam turbines could drive the compressors. Steam turbines are historically the most popular means of driving centrifugal compressors. They have the ability to operate over a wide speed range. Electric motors have experienced increasing favor due to a typically lower operating cost. Buying electricity is more economical than small scale steam generation for a specific piece of equipment (Brown). However, with the desulfurization processes generating steam and with steam available from the power plant, a steam turbine may be the best means of driving the compressors.

Air Compressor Costs

Compressor costs were determined from a budgetary quotation obtained from Ingersoll-Rand. Ingersoll-Rand stated a cost of $241,000 for the Centac Model 2CV23M3EEPF. This model Centac is a centrifugal air compressor (drive and motor) capable of raising 2,250 acfm to

280 psia. Extrapolation was used to determine the cost of compressors needed for the different flow rates. Figures in Peters and Timmerhaus (1991) were used to determine the rate at which compressor costs change with varying flow rates.

The compressor, PRESAIR, is modeled as a six stage compressor. It has been assumed that the interstage coolers lower the air temperature to 115^{o}C. Calculation of stage efficiency was performed using a procedure outlined in Brown (1986). The polytropic efficiencies calculated range from 0.65 to 0.787, which are consistent with other values found in literature (Brown; Dimoplon). PRESAIR pressurizes 8,800 acfm (in the DSRP base case); for such a flow ASPEN predicts a 3,280 HP power requirement. Directly scaling up the Centac (2,250 acfm, 800 HP) compressor predicts a 3,130 HP power requirement. The similar horsepower requirements suggest that ASPEN is realistically simulating the air compressor.

1.d. *RECYCOMP - DSRP*

The compressor RECYCOMP repressurizes the vapor stream leaving the sulfur condenser (the tailgas of the DSRP reaction) and sends it back to the desulfurization reactor. Recycling this stream eliminates an emissions stream while causing a minor load increase for the reactor network. The pressure increase between the condenser and the desulfurization reactor should be within the capabilities of a single stage centrifugal compressor, and RECYCOMP was modeled as such.

1.e. *High Pressure Condenser - DSRP*

The High Pressure Condenser condenses sulfur out of the DSRP Reactor effluent stream. It is high pressure in the sense that it operates near the pressure of the DSRP Reactor. Reducing the temperature to 140^{o}C condenses the sulfur. At this temperature, the vast majority of sulfur condenses, and there is no risk of freezing.

The High Pressure Condenser is simulated using two blocks (Figure 2). The first, COND-I, is an equilibrium block that establishes equilibrium between S_2 and S_8. At high temperatures like those in the DSRP reactor, sulfur is predominately in the S_2 form (Barnett; Chen; Cotton). At the cooler condensation temperatures, the S_8 and S_6 sulfur species predominate. The second block, COND-II, establishes equilibrium between the S_8 and S_6 sulfur species and phase equilibrium. The S_8 and S_6 sulfur species are easier to condense. Calculation of the sulfur equilibrium, in addition to more accurately simulating the phase equilibrium, also increases the accuracy of the heat transfer requirements. The low temperature in the condenser makes it unsuitable for the direct production of high pressure steam. The condenser could be used to preheat the feedwater to other steam-generation units (Appendix F).

1.f. *VAPORIZR - DSRP*

Reducing the sulfur product stream's pressure to ambient will cause the water present in the stream to vaporize. The vaporizing water can cool the sulfur stream enough to cause freezing. The VAPORIZR accomplishes three tasks: a) it reduces sulfur pressure to ambient; b) it supplies heat to the sulfur stream so that the temperature will be maintained at 140^{o}C and sulfur will remain molten; and, c) it also helps purify the product stream by removing water from the sulfur.

1.g. *PD-COOLR - DSRP*

Prior to entering the condenser, the DSRP Reactor effluent ("RXNPRD") is sent through the Product Cooler (PD-COOLR) heat exchanger. Cooling the reactor products in this heat exchanger reduces the condenser heat duty and PD-COOLR operates at temperatures suitable for generating high pressure steam. Sulfur condensation inside the PD-COOLR should be avoided. Condensation would create the undesirable situation of two phase flow and would require removing the sulfur during shutdown so that it will not freeze inside the heat exchanger. Operating the PD-COOLR above the product stream's dew point would prevent sulfur condensation. Dew point calculations were made for the various reactor effluent distributions.

The allotropic sulfur distribution (S_2, S_6, S_8) changes with temperature, however the speed at which equilibrium is reached is unknown. It is not known how closely sulfur allotrope distribution will approach equilibrium in the cooler. Therefore, calculations were made for the dew point temperatures at both the equilibrium distribution of sulfur allotropes, and at the allotrope distribution that leaves the reactor (Table 6).

For the simulations, the PD-COOLR was defined to cool reaction products to 415^{o}C. Table 6 shows that at 415^{o}C sulfur condensation will not occur if the sulfur allotrope equilibrium is reached instantaneously (Sulfur Equilibrium = yes) and also will not occur if the sulfur allotrope distribution is still at the DSRP Reactor temperature distribution (Sulfur Equilibrium = no).

Table 6: Dew Point Temperatures for DSRP Product Distributions

Product distribution	Sulfur Equilibrium	Pressure (psia)	Temperature (^{o}C)
DSRP	yes	275	360
DSRP	no	275	405
DSRP-b	yes	275	357
DSRP-b	no	275	402
DSRP-c	yes	275	362
DSRP-c	no	275	406

1.h. *AIR-HX - DSRP*

The AIR-HX heat exchanger utilizes the hot regenerator off gas (“ROG”) stream to raise the temperature of the high pressure air stream (“P-O2-N2”). Heating the air is required to achieve a sufficiently high temperature to initiate the regeneration reaction. Cooling the ROG reduces the heat removal required to keep the DSRP reactor at 600^{o}C. The hot (above 800^{o}C) ROG stream contains SO_2. The presence of hot SO_2 requires that the AIR-HX heat exchanger tubes be constructed from type 310 stainless steel (SS 310).

2. AHGP Equipment

The AHGP consists of a desulfurization transport reactor and a 3-stage bubbling bed regeneration reactor. The reactions that remove sulfur from coal gas (Reactions 5 and 6) proceed in the desulfurization reactor. In the regenerator the sorbent is regenerated with SO_2, to generate elemental sulfur (reaction 7), and is subsequently regenerated with O_2 to produce SO_2 (reactions 8 & 9). Forming elemental sulfur during regeneration eliminates the need a for third reactor, as the DSRP based process requires. Other major pieces of equipment in the AHGP include compressors, condensers, a demister, and heat exchangers.

2.a. *Desulfurization and Regeneration Reactors - AHGP*

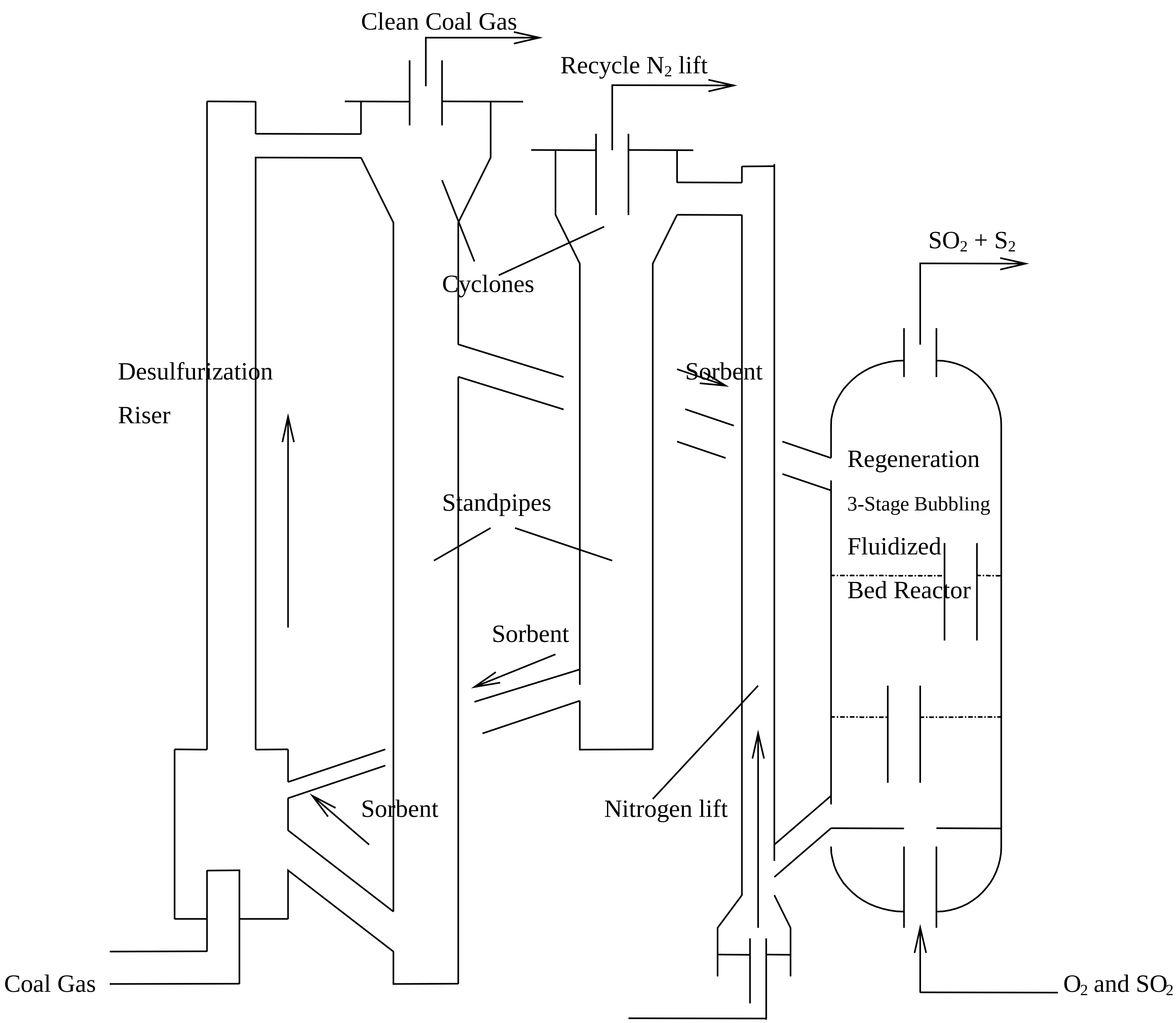
There are several differences between the AHGP desulfurization and regenerator reactor designs (Figure 6) and those envisioned for the DSRP -based process (Figure 5). For example, in the AHGP sorbent descends counter-currently against the rising SO_2 in the regeneration reactor. Sorbent descending through the regenerator makes it necessary to re-elevate sorbent into a standpipe located upstream of the desulfurization reactor. A heat exchanger in the standpipe enables cooling of the sorbent before it re-enters the desulfurization reactor.

The top stage of the regenerator (HX-STAGE, Figure 3) heats the entering sorbent by direct contact with the exiting SO_2 stream. The second stage of the regenerator is modeled with REGEN2 and S-REGEN2. REGEN2 models the following equilibrium reaction:

$$3\ SO_2 + 4\ FeS = 7/2\ S_2 + 2\ Fe_2O_3 \qquad (7)$$

This equilibrium reaction is modeled with an RSTOICH block, assuming a 0.667 fractional conversion of FeS. An RSTOICH block is used due to the difficulty of balancing SO_2 consumption and generation. As discussed earlier in the report (Section II.4), assuming a 0.667

Figure 6: Schematic of AHGP Desulfurization and Regeneration Reactors



fractional conversion may be an optimistically high assumption. If so, more Fe will need to be circulated to make up for the discrepancy. The S-REGEN2 block establishes the equilibrium distribution of sulfur allotropes.

The bottom stage is modeled with the REGEN1 and S-REGEN1 blocks. Oxygen feed to REGEN1 oxidizes the sorbent. Although there is SO_2 present in large quantities in REGEN1, it is assumed not to oxidize any sorbent. Equilibrium conversion for SO_2 oxidation is assumed to be reached in the second stage. Any unreacted FeS present in the sorbent coming from the second stage is expected to react very quickly with oxygen present (reactions 17 & 18). The ZnS is expected to regenerate less rapidly than the iron compound. Uncondensed sulfur recycling back to REGEN1 will quickly oxidize. These reactions are modeled to occur in the following order:

$$S_8 + 8\ O_2 \dashrightarrow 8\ SO_2 \qquad (17)$$

$$S_6 + 6\ O_2 \dashrightarrow 6\ SO_2 \qquad (18)$$

$$2\ FeS + 3.5\ O_2 \dashrightarrow Fe_2O_3 + 2\ SO_2 \qquad (8)$$

$$ZnS + 1.5\ O_2 \dashrightarrow ZnO + SO_2 \qquad (9)$$

The bottom stage is simulated to operate with all oxygen being consumed in REGEN1, and a small portion of ZnS remaining unoxidized.

More than one regeneration reactor maybe used in parallel for the AHGP. Sizing the reactor (Appendix I) revealed that to achieve the desired superficial velocity for removing the larger sulfur quantities requires undesirably large reactor diameters (25+ ft). The larger reactor diameters will require thicker reactor walls (4.5+ in) to contain the high pressures. Reactors in parallel reduce reactor diameter and the required wall thickness resulting in less steel required. A maximum reactor diameter of 13 feet was the guideline used during sizing. The 3-stage regenerator heights were set at 45 feet. It is expected that 5 ft will be needed for the heat exchanging stage, 10 ft for the middle stage, and 2.5 ft for the bottom stage. The rest of the reactor height will be used for phase separation.

The equations used for sizing and costing the AHGP desulfurization and regeneration transport reactor system are described in Appendix G-Calculation of Reactor Size. The actual calculations can be found in Appendix I-Sizing Reactors for the AHGP.

2.b. *LIFTCOMP - AHGP*

The AHGP desulfurization - regeneration transport reactor system requires a means of elevating the sorbent exiting the regeneration reactor. This will be accomplished using a nitrogen lift (Figure 3 and Figure 6). LIFTCOMP increases the pressure of the nitrogen recycle before it enters the nitrogen lift. A cyclone and filters placed upstream of LIFTCOMP and N2-COOLR will prevent sorbent from damaging the compressor.

2.c. *SO2-COMP - AHGP*

SO2-COMP recompresses the SO_2 loop. It is advantageous to recompress the SO_2 loop after the condenser because the lower gas temperature will increase the compressor efficiency and reduce wear on the compressor. The pressure increase required will be obtainable using a single stage centrifugal compressor.

2.d. *CON-COMP - AHGP*

The CON-COMP compressor is used to reintroduce the SO_2 that vaporizes when the sulfur stream is reduced to ambient pressures (LP-COND, Figure 3). The small flow rate means a single stage reciprocating compressor can be used to pressure the SO_2 stream. The pulsing flow of SO_2 coming from CON-COMP will not have a significant effect on the large SO_2 loop.

2.e. *COND-EQ - AHGP*

The condenser, COND-EQ, cools down the SO_2 loop so that sulfur can be condensed out. The stream temperature is reduced to 140^{o}C, and sulfur distribution is established in COND-EQ. It was initially intended that sulfur equilibrium would be calculated using a REQUIL block; however, this caused convergence problems. Using the RSTOIC block eliminates the convergence problem and does not compromise the validity of the results. The sulfur equilibrium distribution was determined in a separate simulation.

$$4\ S_2 \dashrightarrow S_8 \qquad X_{S2} = 0.98 \qquad (15)$$

$$3\ S_2 \dashrightarrow S_6 \qquad X_{S2} = 0.02 \qquad (16)$$

The large vapor stream containing a small volume of molten sulfur will make a demister necessary to isolate the small liquid flow.

2.f. *DEMISTR - AHGP*

The large gas stream of SO_2 will suspend the relatively small flow of condensed sulfur. The demister (DEMISTR) will be necessary for collecting the sulfur. The liquid sulfur accounts for 8 wt% of the stream ("IN-COND"), but only 0.1 vol% of the SO_2 - sulfur flow.

2.g. *LP-COND - AHGP*

Sulfur leaving the demister needs to be brought to ambient pressure for storage. This can be accomplished in a flash tank (LP-COND, Figure 3). The pressure drop vaporizes much of the SO_2 that co-condenses with the sulfur. The temperature drop caused by SO_2 vaporization is not enough to freeze the sulfur. Vaporizing off the SO_2 decreases the sulfur stream temperature to 127^{o}C, well above the melting temperature of sulfur (114^{o}C). The volumetric flow of SO_2 vaporized is 47 times larger than the condensed sulfur flow. The tank should contain a demister pad or some other separation device to prevent sulfur from being entrained with the SO_2 vaporized.

2.h. *HEATX - AHGP*

The HEATX heat exchanger transfers heat from the warm regenerator effluent (SO_2 and sulfur) to preheat the cool regenerator feed stream of recycled SO_2 and oxygen. Sulfur condensation in the heat exchanger should be avoided. If sulfur condenses, the system would have to handle two phase flow from HEATX to the condenser. Shutdown procedures would also require removing sulfur from the heat exchanger to prevent sulfur from freezing inside. Assuming the sulfur allotrope distribution is at equilibrium when condensation occurs, the SO_2 - sulfur stream's dew point is 310^{o}C. Cooling the SO_2 - sulfur stream to no lower than 315^{o}C should prevent condensation from occurring.

2.i. *N2-COOLR - AHGP*

The N2-COOLR cools the nitrogen stream prior to its recompression in LIFTCOMP. Cooling the stream decreases the power required for recompression and reduces the possibility of damaging the compressor. The cool nitrogen stream contributes to reducing the temperature of sorbent feed to the desulfurization reactor. Sorbent entering the compressor would cause damage. Therefore, filters should be installed upstream of the compressor. The filters will also be placed upstream of the heat exchanger (N2-COOLR) to prevent build up of sorbent in the heat exchanger.

2.j. *RCYHEATR - AHGP*

The RCYHEATR was incorporated to ensure that the SO_2 - oxygen feed to the regenerator would be hot enough to initiate the regeneration reactions. Superheated steam is used to raise the SO_2 - oxygen stream temperature, as the separate steam generation process flow sheets show (Appendix F). RCYHEATR works with the HEATX heat exchanger to raise the SO_2 - oxygen stream temperature above 400^{o}C. The RCYHEATR is needed because, HEATX heat transfer is limited to insure no condensation occurs upstream of the condenser.

V. PARAMETRIC STUDIES

Parametric studies were performed to determine how HGD requirements were affected by various coal gas feeds. Inlet H_2S concentrations were varied to simulate variation in sulfur content with different types of coal. Therefore, H_2S concentrations will vary between plants using different coal sources. The effect of power generation capacity was also simulated. Finally, different oxygen sources (air vs. pure oxygen) were investigated. Flow sheets and stream summaries for variations of both processes can be found in appendix H.

1. H_2S Inlet Concentration

DSRP and AHGP simulations were performed using a base case coal gas feed containing 0.85 mol% H_2S and a base case power production of 260 megawatts, after sulfur removal. Additional simulations were performed to determine the effect of H_2S inlet concentration on the amount of coal gas that had to be produced. Table 7 shows how varying H_2S inlet concentration requires increasing the gasification of coal to maintain 260 MW generation.

Table 7: Coal Gas Fed to and Consumed by HGD for Various H_2S Concentrations

Simulation	H_2S inlet conc. (mol%)	Coal Gas Fed (lb/hr)	Consumed H_2 (lb/hr)	Consumed CO (lb/hr)
DSRP	0.85	460,000	320	6,000
DSRP-b	2.50	501,000	1,000	19,000
DSRP-c	0.25	447,000	90	1,700
AHGP	0.85	450,000	160	0
AHGP-b	2.50	468,000	470	0
AHGP-c	0.25	444,000	46	0

The sulfur concentration has a profound effect on DSRP flow requirements because of the coal gas slipstream used in the DSRP reactor. The coal gas slipstream increases as the amount of sulfur converted in the DSRP reactor increases.

The small increase in required coal gas for the AHGP can be attributed to the consumption of H_2 in the desulfurization reaction:

$$Fe_2O_3 + 2\,H_2S + H_2 \dashrightarrow 2\,FeS + 3\,H_2O \qquad (6)$$

The higher sulfur concentrations also require more sorbent circulation to dissipate the heat evolved during reactions. Increased sulfur concentrations require larger reactors. Increasing sulfur also increases the heat removal requirements.

2. Power Generation

Parametric studies were performed to determine the influence of power plant capacity; power generation is 260 MW in the base case. Inlet flows were altered to generate 110 MW and 540 MW. The power level adjustments resulted in flow rates and energy transfer that both scale directly with the change in power generation. The effect of the varying coal gas feed rate was similar to the effect of changing H_2S feed concentrations. An economic comparison shows that the process costs depend on the total sulfur removal requirements. Variations in the flow rates of the other coal gas components do not have a significant effect on the HGD.

3. Pure Oxygen vs. Air Oxidation

Sulfur is removed from the coal gas stream by the reaction of H_2S with the active components of the sorbent to form metal sulfides. Regenerating the sorbent allows it to be reused for removing more sulfur. Sorbent regeneration occurs by exposing the sulfurized sorbent to an oxidizing environment. Pure oxygen and air are both capable of performing the oxidation. Implications of using oxygen and air follow.

3.a. *DSRP*

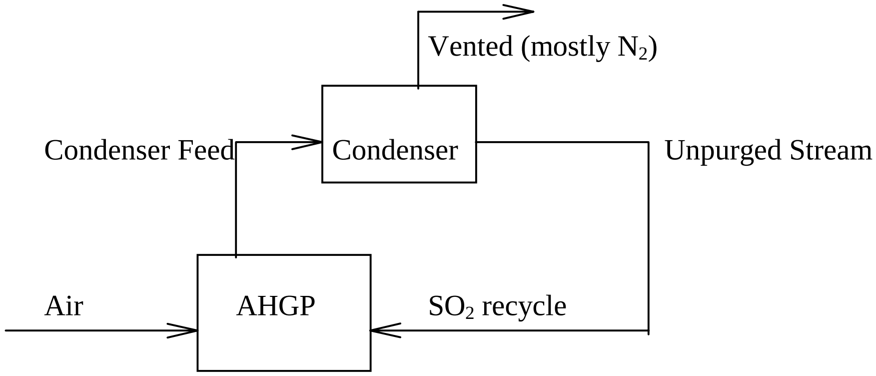
Pure oxygen is an impractical oxidizing medium for sorbent regeneration. In the DSRP - based process, regenerating with pure oxygen would result in such high temperatures that the sorbent would sinter. By comparison, the nitrogen present in air dilutes the oxygen and serves as a heat sink for the highly exothermic regeneration reactions. What is not intuitively obvious is that it is more expensive to supply air to the system than to supply oxygen. For DSRP - based process conditions it is more expensive to compress air than to separate oxygen and then compress only the oxygen (Hvizdos).

3.b. *AHGP*

Air is not a viable oxidizing medium for use in the Advanced Hot Gas Desulfurization Process. The use of air would require separating nitrogen from sulfur dioxide. The AHGP process has a large SO_2 stream that circulates through the regeneration reactor and the sulfur condenser. In the AHGP, oxygen enters the SO_2 loop as a pure oxygen feed and leaves with the sorbent. Sulfur enters the SO_2 loop on the sorbent and leaves as condensed sulfur. Feeding air instead of oxygen would provide a steady flow of nitrogen into the SO_2 loop. Maintaining steady state would require removing nitrogen at the rate it is introduced.

The concept of adding a condenser to the SO_2 loop was investigated for separating nitrogen from SO_2 (Figure 7). ASPEN simulations were performed to determine the condenser conditions necessary for removing nitrogen at the rate it enters the system. The idea was to condense the SO_2 in the loop and vent only nitrogen. Table 8 shows that this concept is impractical. When the ratio of SO_2: N_2 is large the SO_2 is more prone to condense. This can be seen in table 8 where for the same temperature and pressure, uncondensed SO_2 (SO_2 vented) decreases as the mass fraction of SO_2 increases. Therefore, the most efficient condenser will have the minimum amount of N_2 feed to it. The minimum N_2 fed to the condenser will be equal to the rate at which nitrogen enters the system via the air steam. The minimum corresponds to a case where no N_2 condenses (N_2 unpurged). Table 8 shows that even with the very low N_2 concentration there is an unreasonable amount of SO_2 vented.

Figure 7: Condenser for Removal of Nitrogen



The simulations assumed that the total SO_2 loop flow would be 260,000 lbs/hr and 13,500 lbs N_2/hr would need to be removed.

Table 8: N_2 Removal at Various N_2 Concentrations, Condenser Temperatures and Pressures

Condenser Fed: SO_2 mass fraction	Condenser Pressure (psia)	Condenser Temperature (°C)	N_2 unpurged (lbs/hr)	SO_2 vented (lbs/hr)	N_2 vented (lbs/hr)
0.100	275	50	0	26,000	234,000
0.900	275	50	418	58,200	25,600
0.946	275	50	511	30,800	13,500
0.946	400	50	1,010	16,800	13,000
0.940	275	-20	716	1,540	14,900

Furthermore, nitrogen is not needed as a heat sink in the AHGP. The SO_2 stream is a sufficient gas phase heat sink to carry away the heat of the regeneration reaction. The economic analysis showed it is actually desirable to feed oxygen instead of air. The cost of compressing air is higher than the cost of separating out oxygen and then compressing only the oxygen.

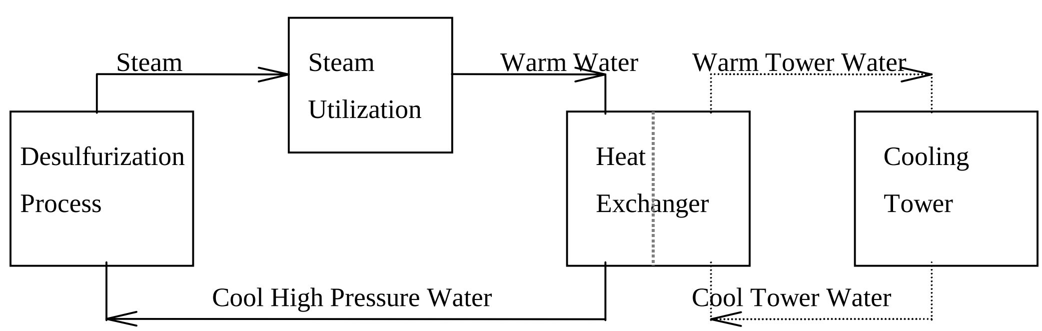
VI. ADDITIONAL PROCESS CONSIDERATIONS

1. Steam Generation

The coal gas desulfurization with sulfur production overall process is exothermic. DSRP and AHGP both require heat removal for condensation and to maintain reaction temperatures. The heat removal requirements create the opportunity to generate high pressure steam that could drive plant equipment or be incorporated into the plant's power generation steam cycle.

Steam generation has been modeled as a closed loop. Steam is generated by removing heat from the desulfurization process. The steam is then utilized, by undefined means, condensed, cooled and the condensate is reused. Cooling tower water is used to cool the steam-condensate loop (Figure 8). There are benefits to having a self-contained loop for steam production. First, it makes it easy to maintain steam-condensate purity, which reduces fouling and corrosion. It also allows for higher cool water feed temperatures (~ 90^{o}C), which increases steam production.

Figure 8: Schematic for HGD Steam Generation



The steam generated from the HGD process was assumed to be at 950 psia and 441^{o}C (Appendix F). Since desulfurization would be incorporated into a larger power generating plant, it is not possible to discern the most useful steam conditions without knowledge of the power generation facility. It is likely that steam generated from the HGD would be utilized by existing power plant equipment. Since the end use of the steam generated is unknown a generic dollar credit for the steam generated was used for the economic analysis. Peters and Timmerhaus (1991) state that 500 psig steam was worth $ 0.0039/lb in 1990; this value was used during the economic assessment. The benefit calculated should be a conservative value since the simulated steam produced is at a higher pressure (950 psia) and the economic calculations use 1996 as a basis. However, another source notes that for 900 psi and 441^{o}C steam, 1 kWh power generation can be expected per 22.44 pounds of steam (Noyes). The economic credit from the conversion of steam to power according to this relationship was less than the credit obtained using the Peters and Timmerhaus relationship. Since the Peters and Timmerhaus credit value is conservative and still predicts a larger benefit, the Peters and Timmerhaus value was used.

2. Material of Construction

Type 310 stainless steel (25%Cr - 20%Ni) should be used for the construction of equipment that contacts sulfur species. Type 310 stainless steel (SS 310) will be more durable than type 316 stainless steel (SS 316) (17%Cr - 8% Ni - 2%Mo). Higher chromium content gives SS310 greater oxidation resistance, and the higher nickel concentration gives improved resistance to carburization (EPRI). Cost data for SS310 is not contained in ASPEN so SS316 material cost factors were used.

3. Sulfur Storage

Transporting molten sulfur is preferred over solid sulfur. Liquid sulfur is easier to transport and reduces handling losses. It will be necessary to store the molten sulfur before it is shipped out by train. The storage tank should be capable of storing several days worth of recovered sulfur. It should also be equipped with a heat exchanger to keep sulfur molten. The costs of the sulfur storage tanks were calculated using ASPEN assuming SS 310 was used to construct storage for seven days of sulfur production (SS 316 was entered in ASPEN due to lack of data for SS 310).

4. Process Operation

The DSRP should be the easier process to operate. Balancing the SO_2 production and consumption in the AHGP appears to be particularly difficult. The difficulty arises from the reaction of FeS with SO_2 to form elemental sulfur. The reaction's equilibrium varies significantly with temperature. If the reactants are too thermodynamically favored, less SO_2 will be consumed than expected. However, SO_2 production will remain constant (sorbent oxidation being driven to completion by oxygen). Thus, if the reaction:

$$3\ SO_2 + 4\ FeS \Longleftrightarrow 7/2\ S_2 + 2\ Fe_2O_3 \qquad (7)$$

does not reach design conversions, SO_2 flow will increase and sulfurized sorbent will be returned to the desulfurization reactor. With SO_2 already present in great excess the increased SO_2 flow will not significantly shift equilibrium towards the products.

It is recommended that the AHGP be operated at conditions that will cause a net consumption of SO_2. Replenishing depleted SO_2 levels can easily be accomplished by increasing the oxygen feed. Excess oxygen will convert elemental sulfur into SO_2.

Preventing the build up of impurities in the SO_2 loop contributes to the complexity of the AHGP. Venting a portion of the loop is undesirable since it contains mostly SO_2. Venting would release SO_2, emissions the system is designed to eliminate. Operating the AHGP requires determining the rate at which impurities build up in the recycle loop and the appropriate purge stream for the rate of build up. The purge stream should be fed to the desulfurization reactor, reducing the release of SO_2.

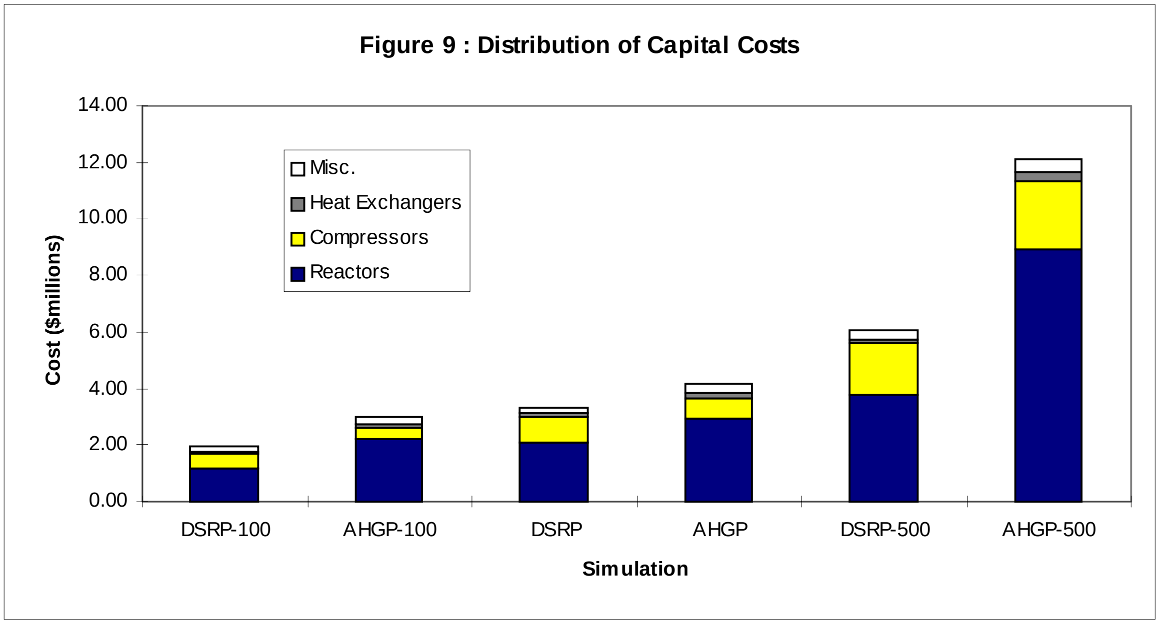
VII. ECONOMIC ANALYSIS

1. Capital Expenditures

The AHGP requires more capital investment than the DSRP. Reactors account for over half of the capital investment. The higher cost of AHGP reactors results in an higher overall capital investment necessary for the AHGP (Figure 9). The majority of equipment was costed using ASPEN. Equipment costed by ASPEN has a purchase date set at June, 1996. Equipment contacting sulfur will experience less corrosion when constructed of stainless steel 310 (SS310). Since ASPEN lacks material of construction correction factors for SS310, SS316 values were used. While the majority of equipment was costed using ASPEN, the equipment that comprises the majority of the capital expenditures, such as the reactors, were estimated by other means.

The reactor costs were calculated using a procedure outlined in Peters and Timmerhaus (1991). The reactor costs were determined using the amount of steel required for their construction. The procedure is described in appendix G, and the calculations are contained in appendix H and appendix I. The reactor cost includes the cost of installation.

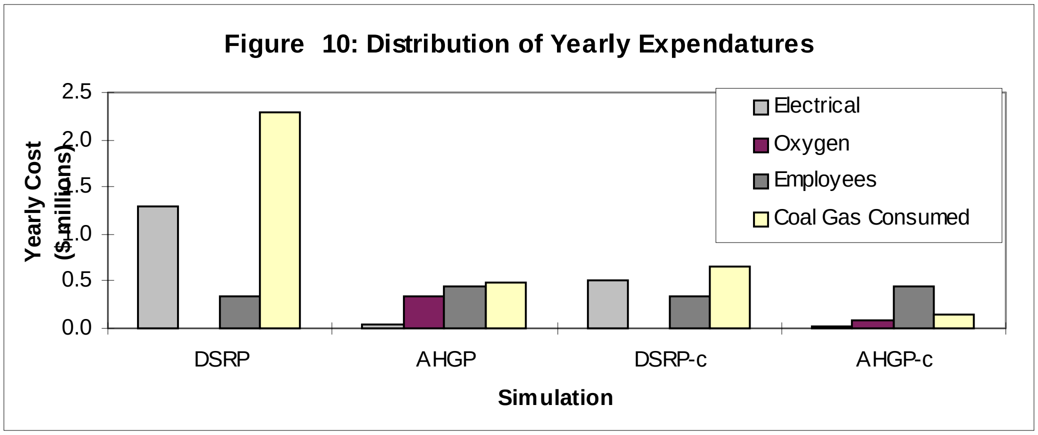
Another piece of equipment not costed by ASPEN is the PRESAIR - air compressor used in the DSRP. PRESAIR costs were determined by scaling a price quote for the Ingersoll-Rand Centac air compressor. The Centac Model 2CV23M3EEPF, capable of raising 2,250 acfm to 280 psia, was quoted at $241,000. Extrapolation was used in determining the cost of compressors needed for the different flow rates. Figures in Peters and Timmerhaus (1990) were used to determine the rate at which compressor costs change with varying flow rates.

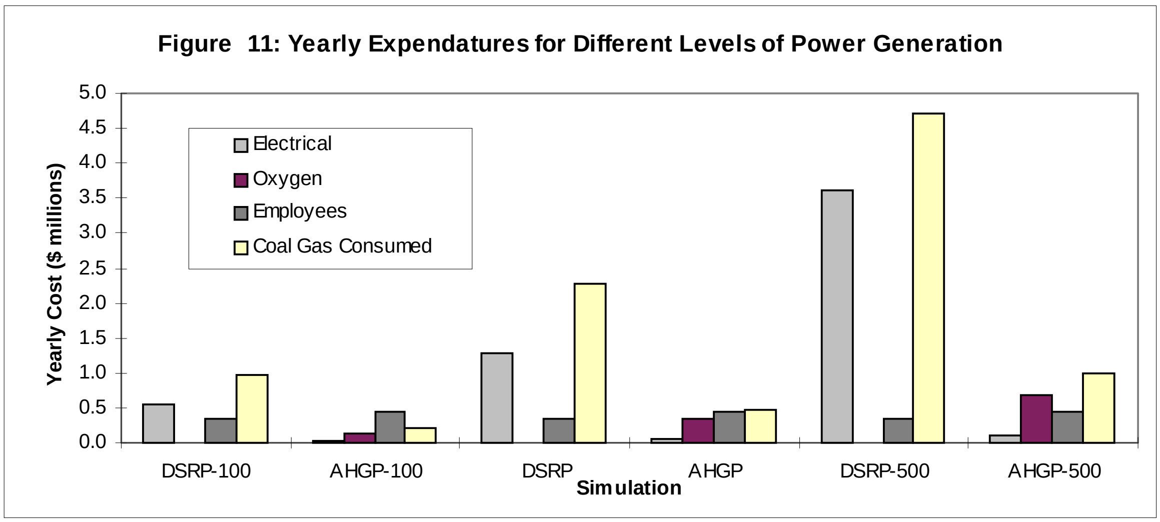


There are additional capital costs not included in this report, two of which, piping costs and sorbent/catalyst costs, will probably be significant. There will be other expenses, like additional office space for employees, which are site dependent. The site dependent expenses should not have an significant effect on the total capital investment calculations. At this stage of investigation the piping and sorbent/catalyst cost are assumed identical for both HGD process. If this assumption is valid than a comparison of the overall capital costs for the AHGP and the DSRP will not be affected by their absence.

2. Yearly Operating Costs

The AHGP has a lower yearly operating cost than the DSRP. Figures 10 and 11 show the distribution of the major yearly expenditures for both processes.





The bases cases (DSRP and AHGP) have coal gas feeds containing 0.85 mol% H_2S and produce 260 MW. Most of the yearly expenditures decline as the amount sulfur in the coal gas is decreased (DSRP-c and AHGP-c have feeds containing 0.25 mol% H_2S). The exception is the yearly costs of additional employees, which have been assumed to be dependent on the complexity of the HGD process and not its size. As the sulfur concentration decreases both the absolute expenditure difference (DSRP cost - AHGP cost) and the relative expenditure difference ([DSRP cost - AHGP cost] / AHGP cost) decrease. This decrease indicates that the competitive advantage of the AHGP is smaller for cleaning a coal gas stream containing a low H_2S concentration. The same trend exists comparing the economics of different levels of power generation: the AHGP's yearly economic advantage over the DSRP declines as the overall power generation is decreased.

Figure 12: H2S Concentration's Effect on HGD Yearly Operating Costs

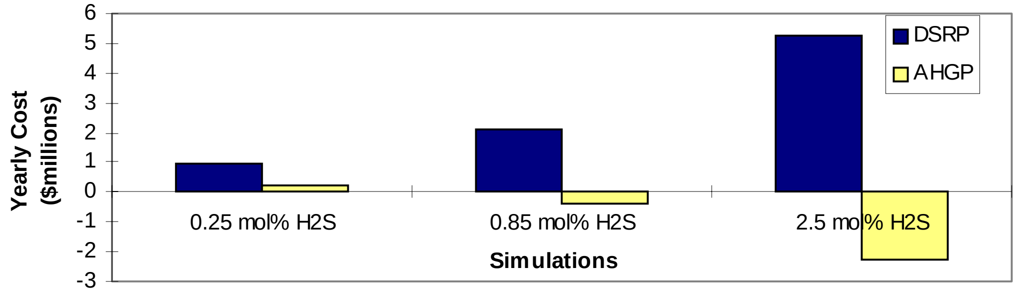
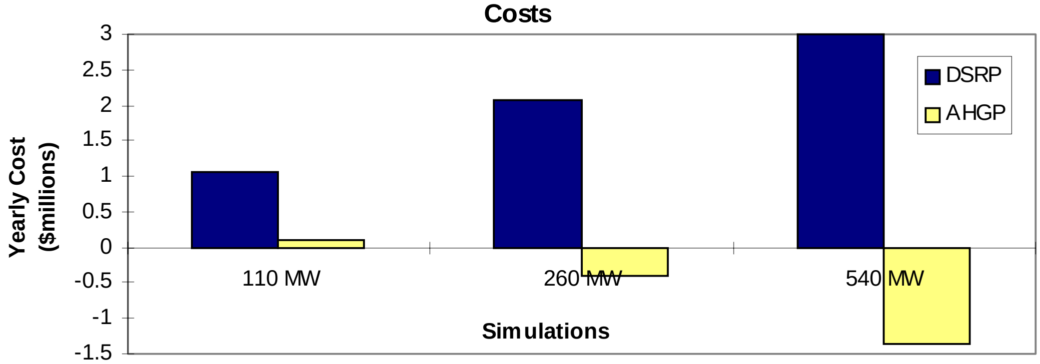
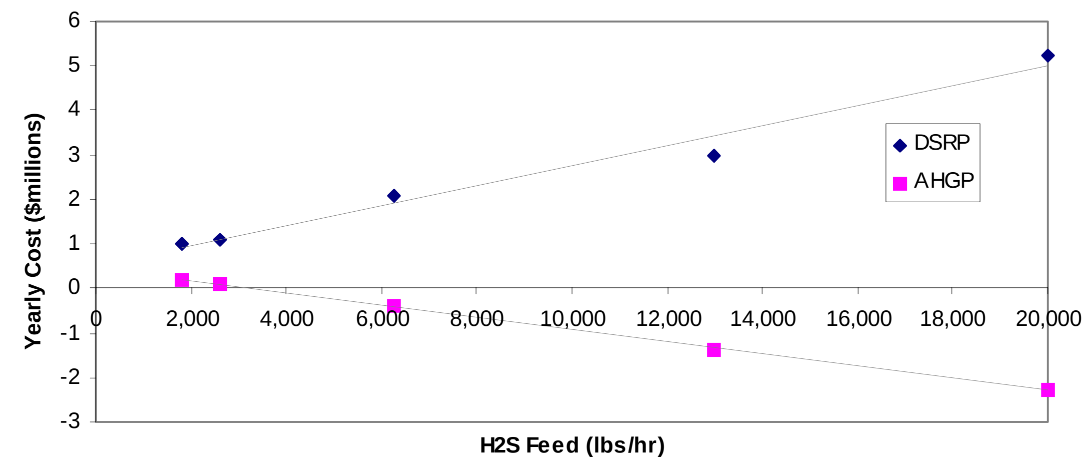


Figure 13: Power Generation's Effect on HGD Yearly Operating Costs



In assessing the yearly cost of maintaining HGD, benefits of the process should also be accounted for. Two sources of credit were observed: the recovery of sulfur and the production of steam. Sulfur credits where consistently larger than steam credits within the same simulation. The sulfur credits remained virtually unchanged between corresponding DSRP and AHGP simulations. Figure 12 and 13 show that for several AHGP conditions the credits are larger than the expenditures. This results in negative yearly operating costs. When larger amounts of sulfur are removed, the yearly expenditures combined with the sulfur and steam credits result in negative yearly costs for the AHGP. In such cases it is more profitable to use the AHGP, then to leave the coal gas stream untreated (if Federal Regulations allowed). The profit that results from the sale of recovered sulfur (Appendix M) allows the AHGP to be more profitable than generating power without desulfurization.

Figure 14: Yearly Costs as a Function of Sulfur Feed



The yearly costs have a linear dependence on the amount of sulfur being processed. This can be seen by comparing all simulations (DSRP, DSRP-b, DSRP-c, DSRP-100, DSRP-500, AHGP, AHGP-b, AHGP-c, AHGP-100, AHGP-500). Figure 14 shows that regardless of how the sulfur feed is varied (changing concentration vs. changing power generation), the yearly costs scale directly with sulfur removed.

2.a. *Electrical*

The pumps and compressors have been assumed to account for the majority of the electrical requirements for the HGD processes. The additional power requirements for lighting and instrumentation have been assumed to be 20% of the compressor and pump requirements for the base case of each HGD. It is assumed that the additional power requirements will not vary significantly with plant size.

The DSRP power requirement is significantly higher than that of the AHGP. The PRESAIR air compressor is the reason for the high DSRP power requirement. The air compressor supplies air to the regenerator for the oxidation of sulfurized sorbent. It is interesting to note that the cost of supplying oxygen by compressing air is more than the cost of separating oxygen and then compressing the pure oxygen. The phenomenon is not unprecedented; it has been observed that as the pressure of injection is raised the cost of compressing air increases faster than the cost of separating oxygen and pressuring only oxygen (Hvizdos).

The compressed nitrogen feed to the DSRP - based process regenerator that is included in the air stream will increase the total volumetric flow to the turbine. This would indicate that there should be a power credit associated with the nitrogen's introduction, offsetting some of the compression costs. However, nitrogen will also increase the heat capacity of the stream, lowering the combustion temperature, thus lowering the power production. These competing effects have been assumed to cancel each other out. The design work assumes there is no change in power production attributed to the introduction of nitrogen.

2.b. *Cooling Water*

The steam generation/cooling loop is closed; maintaining water purity is not difficult for a self-contained loop. Furthermore, makeup water requirements will be negligible, for the detail level of this report. The is no debit calculated for the HGD steam system water because of the above mentioned reasons.

The steam condensate is assumed to be cooled to 90^{o}C by cooling tower water. Tower water is exposed to the atmosphere, which means maintaining water purity will be an issue. There will also be makeup water requirements. Therefore a yearly debit has been calculated for the use of tower water. The tower water flow rates have been calculated in the Complete Steam Generation Scheme simulations (Appendix F). The tower water cools the steam stream that is considered "utilized." Utilized steam is a stream that was steam (441^{o}C, 950 psia) but has been reduced to 30 psia and the corresponding bubble point temperature. Tower water cools the utilized steam stream to 90^{o}C, before its reuse. The cost of the tower water is \$2.6x10^{-5}/lb (Peters). The cost of the tower water is insignificant compared to the other yearly capital expenditures.

The cost of the tower is not an issue as there will already be a tower on site. HGD water sent to it will represent only a minor increase in load.

2.c. *Oxygen*

The cost of supplying oxygen has been assessed as a yearly expenditure with no capital cost. Dr. George Roberts indicated that its reasonable to expect oxygen to cost \$20/ton. The value is reasonable when compared with a dated guideline (Chilton, 1960) stating 99.5% pure oxygen at 450 psig would sell at \$8 to \$15/ton. There are no capital costs associated with the supplied oxygen assuming the oxygen will be bought from a gas supplier, in which only a usage charge is assigned. The price has been assumed to be set at \$20/ton, the price will actually be dependent on usage. The unit cost of oxygen decreases as quantity purchased increases.

There are oxygen costs only for the AHGP, since air is used to oxidize the sorbent in the DSRP.

2.d. *Additional Employees*

The number of additional employees required to operate the HGD processes have been assumed constant with process size. The additional employees required will depend more upon the complexity of the process than its size. The hiring of two additional engineers and two maintenance personal have been assigned to the DSRP. The AHGP has the hiring of three engineers accounted for. An additional engineer is hired since the AHGP is a more complex process to control because SO_2 production and consumption must be balanced. Furthermore, the purity of the SO_2 loop must be maintained. Two maintenance personnel are also accounted for in AHGP costs. The unit cost for an engineer is assumed to be \$100,000/year, and maintenance personnel are assumed to cost \$70,000/year. These numbers include the base salary and benefits.

2.e. *Consumed Coal Gas*

Coal gas (H_2 and CO) is consumed in both HGD processes. The consumption reduces the amount power that can be produced. The cost of consumed coal gas is calculated from the CO and H_2 lost during HGD, and calculating the value of the energy that the CO and H_2 could have produced. Calculation of power generation is described in Appendix J.

The DSRP consumes substantially more coal gas then the AHGP; this is the major factor in the lower yearly operating cost of the AHGP.

2.f. *Additional Yearly Expenditures*

Sorbent and catalyst attrition have not been accounted for in this report. The rate at which sorbent and catalyst need to be replaced times their unit cost will represent another yearly expenditure. Assuming the attrition costs for both processes are identical a comparison of the process economics will be unaffected by the absence of attrition costs in this report.

Maintenance charges have not be fully accounted for in this report. While the cost of additional employees to maintain equipment has been included, the cost of the replacement parts and equipment have not. Yearly maintenance costs should increase with years of service as well as with the size of the HGD process.

3. Economic Summary

The AHGP has a higher initial startup costs, indicated by its larger capital requirements. However, the AHGP has lower yearly expenditures then the DSRP. The operating cost difference is large enough to offset the initial startup cost difference within a few years.

Figure 15: Cumulative HGD Investment

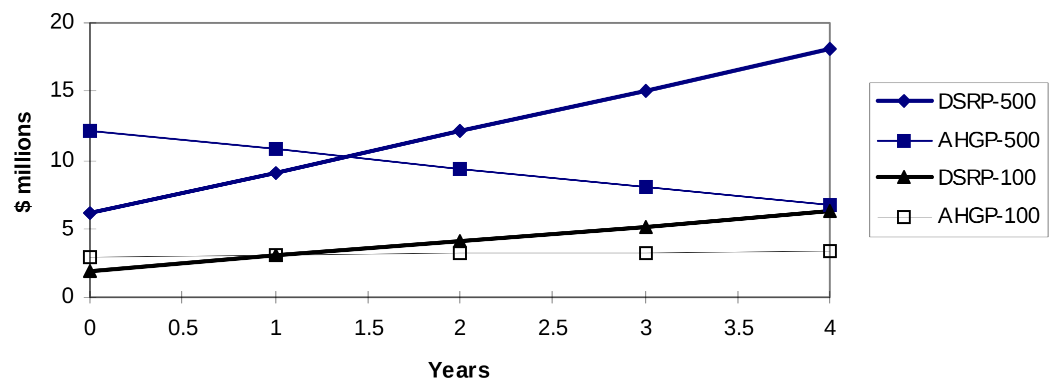


Figure 15 shows that despite an higher initial investment, within two years the AHGP can financially outperform the DSRP.

VIII. SUMMARY

Mass and energy balances were calculated for the Direct Sulfur Recovery Process - based Hot Gas Desulfurization and the Advanced Hot Gas Process. Establishing the balances has helped determine the equipment requirements for both processes. The specifications for the major pieces of equipment have been described in this report.

Simulating the HGD processes revealed the complexity of both processes. The AHGP appears to be the more difficult of the two processes to operate. More employees may be needed to operate the AHGP process than the DSRP -based process.

Capital costs for the AHGP are higher than those for the DSRP. However, yearly operating costs for the AHGP are considerably less than those of the DSRP. After two years of operation the total cost of implementing an AHGP will be less then the cost of a DSRP -based process. It will be more difficult to operate an AHGP but the substantial savings the process delivers makes it the more desirable process to implement.

REFERENCES

ASPEN PLUS, *Reference Manual - Volume 2 Physical Property Methods and Models*, Release 9, Aspen Technology, INC, 1994

Barnett, E. B., Wilson, C. L., *Inorganic Chemistry - A Text-book For Advanced Students*, 2nd Edition, John Wiley & Sons Inc., New York, 1957

Bird, R.B., Stewart, W.E., Lightfoot, E.N., *Transport Phenomena*, John Wiley & Sons, New York, 1960

Bloch, H. P., *A Practical Guide to Compressor Technology*, McGraw-Hill, New York 1996

Braker, W., Mossman, A. L., *Matheson Gas Data Book*, 5th Edition, Matheson Gas Products, Milwaukee, 1971

Brown, R. N., *Compressors - Selection & Sizing*, Gulf Publishing Company, Houston 1986

Campbell, W. M., and Henningsen, G. B., "Hot Gas Desulfurization Using Transport Reactors" *Proceedings of the Twelfth Annual International Pittsburgh Coal Conference*, Pittsburgh, 1995

Caruanan, C. M., "Processes Aim to Improve Sulfur Recovery From Gases," *Chemical Engineering Progress*, February 1996, pp. 11-17

Chen, D. H., *Kinetically Modified Equilibrium Calculation and Thermodynamic Analysis for the DSRP Hot Gas Sulfur Recovery Process*, Final Report to Research Triangle Institute, March 10, 1994 report 1994

Chilton, C. H., Cost Engineering In the Process Industries, McGraw-Hill Company, New York, 1960, pg. 302-303

Cootner, P. H., Lof, G. O., *Water Demand for Steam Electric Generation*, John Hopkins Press, Baltimore, 1965

Cotton, F. S., Wilkinson, G., *Advanced Inorganic Chemistry - A Comprehensive Text*, 2nd Edition, Interscience Publishers, New York 1966

Cusumano, J. A., Dalla Betta, and R. A., Levy R. B., *Catalysis in Coal Conversion*, Academic Press, Inc. New York 1978

Dimoplon, W., "What Process Engineers Need to Know About Compressors," *Hydrocarbon Processing* May 1978, pp. 221-227

EPRI CS-4035, “Continuous Testing of the Resox Process,” Project 1257-2, Final Report, May 1985

Gangwal, S. K., Hot-Gas Desulphurisation Sorbent Development for IGCC Systems, “Desulphurisation 2 Technologies and Strategies for Reducing Sulphur Emissions” *Institution of Chemical Engineers*, New York 1991

Hvizdos, L. J., Howard, J. V., Roberts, G. W., “Enhanced Oil Recovery Through Oxygen Enriched In-Situ Combustion: Test Results From the Forest Hill Field in Texas” *Journal of Petroleum Technology* June 1983, pp. 1061-1070

Jaffee, R. I., “Advances in Materials for Fossil Power Plants,” *Advances in Material Technology for Fossil Power Plant - Proceedings of an International Conference on Advances in Material Technology for Fossil Power Plants,* ASM International, Chicago, 1987

Kern, D. Q., *Process Heat Transfer*, McGraw-Hill Book Company, New York, 1950

Noyes, R. Cogeneration of Steam and Electric Power, Noyes Data Corporation Park Ridge, Park Ridge, NJ, 1978

Perry, R.H., Green, D.W., Maloney, J.O., *Perry’s Chemical Engineers’ Handbook*, 6th Edition, McGraw Hill Inc., New York, 1984

Peters, M. S., Timmerhaus, K. D., *Plant Design and Economics for Chemical Engineers*, 4th Edition, McGraw Hill Inc. New York, 1991

Portzer, J. W., Research Triangle Institute internal memo, November 26, 1996

Ulrich, G. D., A Guide to Chemical Engineering Process Design And Economics, John Wiley & Sons, Inc. New York 1984

U.S. Geological Survey, Minerals Information, *http://minerals.er.usgs.gov/minerals/pubs/commodity/sulfur/stat/*

Welty, J.R., Wicks, C.E., Wilson, R.E., *Fundamentals of Momentum, Heat, and Mass Transfer*, 3rd Edition, John Wiley & Sons, New York, 1984

Appendix A
Calculation of the SO_2 Circulation Rate for AHGP

SO_2 circulation rates are set to create the desired flow conditions in the regenerating reactor. First the sorbent flow rate through the regenerator must be determined. Al_2O_3 must pass through the reactor in large quantities to keep the adiabatic temperature raise small. The sorbent flow is used to determine the reactor's cross sectional area. The SO_2 circulation rate necessary to provide a 2.5 cm/s upwards velocity is then calculated. Calculation results follow:

SO2 Regenerator Sizing - Commercial Embodiment

Givens:	AGHP (SO2 Regen) Case E-2	AHGP-b (SO2 Regen) Case E-2	AHGP-c
Sorbent circulation rate, lb/hr	166010	496000	48000
Sorbent bulk density, lb/ft3	62.4	62.4	62.4
Req'd rxtr residence time, hr	1	1	1
Regen Gas v_{super}, cm/sec	2.5	2.5	2.5
Desired H/D	2	2	2
Adjusted values:			
Assumed Bed Depth, ft	10	10	10
SO2 needed ft3/hr	**79,813**	**238,462**	**23,077**
Calculated values:			
Hold-up volume, ft3	2660	7949	769
Diameter, ft	18	32	9.9
X-section area, ft2	266	795	77
Calculated H/D	0.54	0.31	1.01
RG Vol. flow rate, acf/sec	21.8	65.2	6.3
RG flow rate, lb/hr	86366	258043	24972
Ratio of RG flow/sorbent, lb/lb	0.52	0.52	0.52
Calculated Bed Depth, ft			
Operating conditions/Gas Density Calc'ns:			
Pressure, psig	275	275	275
Pressure, psia	289.7	289.7	289.7
MW of gas	64	64	64
Bed Temp., C	600	600	600
Bed Temp., R	1571.67	1571.67	1571.67
R, gas constant,	10.73	10.73	10.73
Gas density, lb/ft3	1.1	1.1	1.1

Appendix B
Heat Transfer Coefficients

The following approximate overall heat transfer coefficients were found in the literature. The values were in used estimating the heat exchangers' overall heat transfer coefficients.

	Coolers	
Hot Fluid	Cold Fluid	Overall U_D, BTU/hr ft^2 ^oF
Water	Water	250 - 500
Gases	Water	2 - 50

	Heaters	
Hot Fluid	Cold Fluid	Overall U_D, BTU/hr ft^2 ^oF
Steam	Water	200 - 700
Steam	Gases	5 - 50

Values above found in Kern (1950).

Fluid combination	U, BTU/hr ft^2 ^oF
Water to compressed air	10 - 30
Water to water	150-275
Steam to aqueous solutions	100-600
Steam to gases	5 - 50

Values above found in Welty, Wicks, and Wilson (1984).

Appendix C
Determination of Catalyst Velocity in DSRP Reactor

In order to determine whether the catalyst in the DSRP Reactor (a fast fluid-bed reactor) will be transported to the top of the reactor by the gas feed, the following calculation was performed. A terminal velocity calculation was performed on a catalyst particle. This calculation will approximate the catalyst's velocity relative to the gas phase. The gas velocity through the DSRP will be 3 ft/s (0.9 m/s). The catalyst's relative velocity needs to be less than the gas velocity in order for the catalyst to be elevated.

Terminal velocity is determined from a force balance on the particle.

$$m\frac{dv}{dt} = F_g - F_d - F_b$$

$$m\frac{dv}{dt} = mg - \frac{C_D v^2 \rho(\pi D_p^2 / 4)}{2} - \frac{m\rho g}{\rho_p}$$

At steady state the left side equals zero and the equations simplify to give the steady state (terminal) velocity:

$$v_{ss} = \sqrt{\frac{4}{3}(\frac{D_p g}{C_D \rho})(\rho_p - \rho)}$$

The catalyst size is 160 micron.
$D_p = 1.6 \times 10^{-4}$ m $\qquad$ $\rho_p = 1.2$ g/cm^3 $\qquad$ $g = 9.8$ m/s^2

Bulk samples of the catalyst have a density (ρ_{bulk}) of 0.9 g/cm^3. The bulk catalyst is assumed to have a packing fraction of 0.74, the highest packing fraction possible for spheres. Assuming the packing fraction enables calculation of the individual catalyst density (ρ_p).

$\rho_p = \rho_{bulk}$ /(packing fraction) $\qquad$ $\rho_p = 1.2$ g/cm^3 = (0.9 g/cm^3) / (0.74)

The gas density is taken as a weighted average of the feeds ROG-COOL and SLIPSTRM.

ρ = 0.50 lb/ft^3 x (1,000 gr) / (2.205 lb) x (1 ft^3) / (30.48 cm)^3 = 0.008 g/cm^3 = 8 kg/m^3

Inserting the values gives:

$$v_{ss} = \sqrt{\frac{0.3136 \frac{m^2}{s^2}}{C_D}}$$

The drag coefficient C_D is correlated with the Reynolds number (N_{Re})of the gas phase. After determining the Reynolds number C_D can be determined from charts in Bird (1960).

$$N_{Re} = D_p \, v_{ss} \, \rho \, / \, \mu$$

The steady state velocity is determined iteratively. That leaves μ, viscosity of the gas, the only other unknown.

For viscosity calculations, the gas will been assumed to have the properties of nitrogen (N_2 represents over 50 wt% of reactor gas).

Reactor conditions T = 600°C and P = 275 psia.

The Reichenberg correlation was used for the determination of the high pressure viscosity (Perry's 3-279). The correlation typically has errors of less than 10 percent.

<u>Equations</u>

$$(\mu - \mu^o)/(\mu^o \wp) = A \, P_r^{1.5} \, / \, [\, B \, P_r + (1 + C \, P_r^{D})^{-1}]$$

$$A = 1.9824 \times 10^{-3} \, T_r^{-1} \exp(5.2683 \, T_r^{-0.5767})$$

$$B = A \, (1.6552 \, T_r - 1.2760)$$

$$C = 0.1319 \, T_r^{-1} \exp(3.7035 \, T_r^{-79.8678})$$

$$D = 2.9496 \, T_r^{-1} \exp(2.9190 \, T_r^{-16.6169})$$

<u>Nitrogen Properties</u>

$T_r = T / T_c = 873 \text{ K} / 126.2 \text{ K} = 6.91$ $\quad$ $P_r = P / P_c = 275 \text{ psia} / 492 \text{ psia} = 0.559$

$\mu^o = \mu\,(1 \text{ atm}, 873 \text{ K}) = 3.8 \times 10^{8}$ Poise

And for nonpolar molecules: $\wp = 1$

<u>Calculated Values</u>

A = 0.001615 B = 0.0164 C = 0.01909 D = 0.4269

$$(\mu - \mu^o) / (\mu^o) = 6.7498 \times 10^{-4} / 0.9945 = 6.787 \times 10^{-4}$$

$$\mu = 3.8 \times 10^{-4} + (3.8 \times 10^{-4})(6.787 \times 10^{-4}) \text{ Poise}$$

$$\mu = 3.8 \text{ x } 10^{-4} \text{ Poise} = 3.8 \text{ x } 10^{-5} \text{ Pa s} = 3.8 \text{ x } 10^{-5} \text{ kg/(m s)}$$

The Reynolds number is can now be expressed:

$$N_{Re} = (1.6 \text{ x } 10^{-4} \text{ m})\ (v_{ss})\ (8 \text{ kg/m}^3) / [\ 3.8 \text{ x } 10^{-5} \text{ kg/(m s)}]$$

$$N_{Re} = (v_{ss})\ 33.68 \text{ s/m}$$

And our velocity equation is: $v_{SS} = \sqrt{\dfrac{0.3136 \dfrac{m^2}{s^2}}{C_D}}$

First Iteration, take v_{ss} = 0.9 m/s then N_{Re} = 30 (above equation)
For the above Reynolds number C_D = 2.4 (Fig. 6.3-1 in Bird)
Velocity equation gives v_{ss} = 0.36 m/s

The calculations are repeated.

Second iteration:	v_{ss} = 0.36 m/s	N_{Re} = 12
	C_D = 4.2	v_{ss} = 0.27 m/s
Third iteration:	v_{ss} = 0.27 m/s	N_{Re} = 9.09
	C_D = 4.9	v_{ss} = 0.252 m/s
Fourth iteration:	v_{ss} 0.252 m/s	N_{Re} = 8.49
	C_D = 5.13	v_{ss} = 0.247 m/s
Fifth iteration:	v_{ss} 0.247 m/s	N_{Re} = 8.33
	C_D = 5.19	v_{ss} = 0.246 m/s

The velocity of falling catalyst is 0.25 m/s. Thus in a gas stream flowing up at 0.9 m/s the catalyst will rise at 0.65 m/s (2.1 ft/s).

CONCLUSION: The gas stream will be capable of elevating the catalyst.
Sorbent in the risers will be elevated at approximately the same velocity (20 ft/s) as the gas lifting it.

Appendix D
Calculation of DSRP Catalyst Cycling Rate

The rate at with catalyst is fed to the DSRP was determined by the heat removal requirements of the DSRP reactor. Heat is removed from the reactor by cooling the catalyst effluent and reintroducing that catalyst. Exiting catalyst temperature is set at 600^oC and the catalyst is cooled to 500^oC.

Catalyst Properties

The DSRP reactor catalyst is a porous aluminum oxide catalyst modeled as Al_2O_3. Catalyst density at ambient conditions is 56.18 lb/ft^3. This density includes the void space filled by air. ASPEN was utilized to determine the void space in the settled catalyst, assuming nitrogen fills the voids in the solid catalyst. At ambient conditions 1 lb of Al_2O_3 and 0.00095 lb of N_2 have a combined density of 55.6 lb/ft^3.

The similar densities allow us to assume that there is roughly 0.00095 lb of nitrogen present for every 1 lb of solid Al_2O_3 (at ambient conditions). That quantity of nitrogen occupies 0.0132 ft^3 (at ambient conditions). This represents the catalyst void volume and is expected to remain constant.

$V_{void} = 0.0132 \text{ ft}^3 / 1 \text{ lb } Al_2O_3$

The density of the gas in the reactor was taken as the average of nitrogen's density at 600^oC (275 psia) and 500^oC (275 psia).

$$\rho_{gas} = 0.483 \text{ lb/ft}^3$$

Therefore the mass of gas (in the settled catalyst) per pound Al_2O_3 can be calculated.

$$M_{gas} = 0.483 \text{ lb/ft}^3 \times 0.0132 \text{ ft}^3 = 0.0064 \text{ lbs}$$

The heat transfer requirements for cooling Al_2O_3 were than simulated (including cooling nitrogen contained in the catalyst voids).

$$Q(600^oC \rightarrow 500^oC) = -51.239 \text{ BTU/lb } Al_2O_3$$

Calculation of necessary catalyst circulation rate:

$$(\text{circulation rate }\{\text{lb/hr}\}) = (Q_{DSRP}) / (\ -51.239\ \text{BTU/lb } Al_2O_3)$$

DSRP (circulation rate {lb/hr}) = (-15,340,000 BTU/hr) / (-51.239 BTU/lb Al_2O_3)
= 300,000 lb Al_2O_3 / hr

DSRP-b (circulation rate {lb/hr}) = (-51,320,000 BTU/hr) / (-51.239 BTU/lb Al_2O_3)
= 1,000,000 lb Al_2O_3 / hr

DSRP-c (circulation rate {lb/hr}) = (-4,029,000 BTU/hr) / (-51.239 BTU/lb Al_2O_3)
= 79,000 lb Al_2O_3 / hr

DSRP-100 (circulation rate {lb/hr}) = (-6,459,000 BTU/hr) / (-51.239 BTU/lb Al_2O_3)
= 130,000 lb Al_2O_3 / hr

DSRP-500 (circulation rate {lb/hr}) = (-31,370,000 BTU/hr) / (-51.239 BTU/lb Al_2O_3)
= 610,000 lb Al_2O_3 / hr

Appendix E
Process Flowsheets and Stream Summaries

Direct Sulfur Recovery Process Simulations

DSRP (base case)	0.85 mole% H_2S	260 MW generated
DSRP-b	2.50 mole% H_2S	260 MW generated
DSRP-c	0.25 mole% H_2S	260 MW generated
DSRP-100*	0.85 mole% H_2S	110 MW generated
DSRP-500	0.85 mole% H_2S	540 MW generated

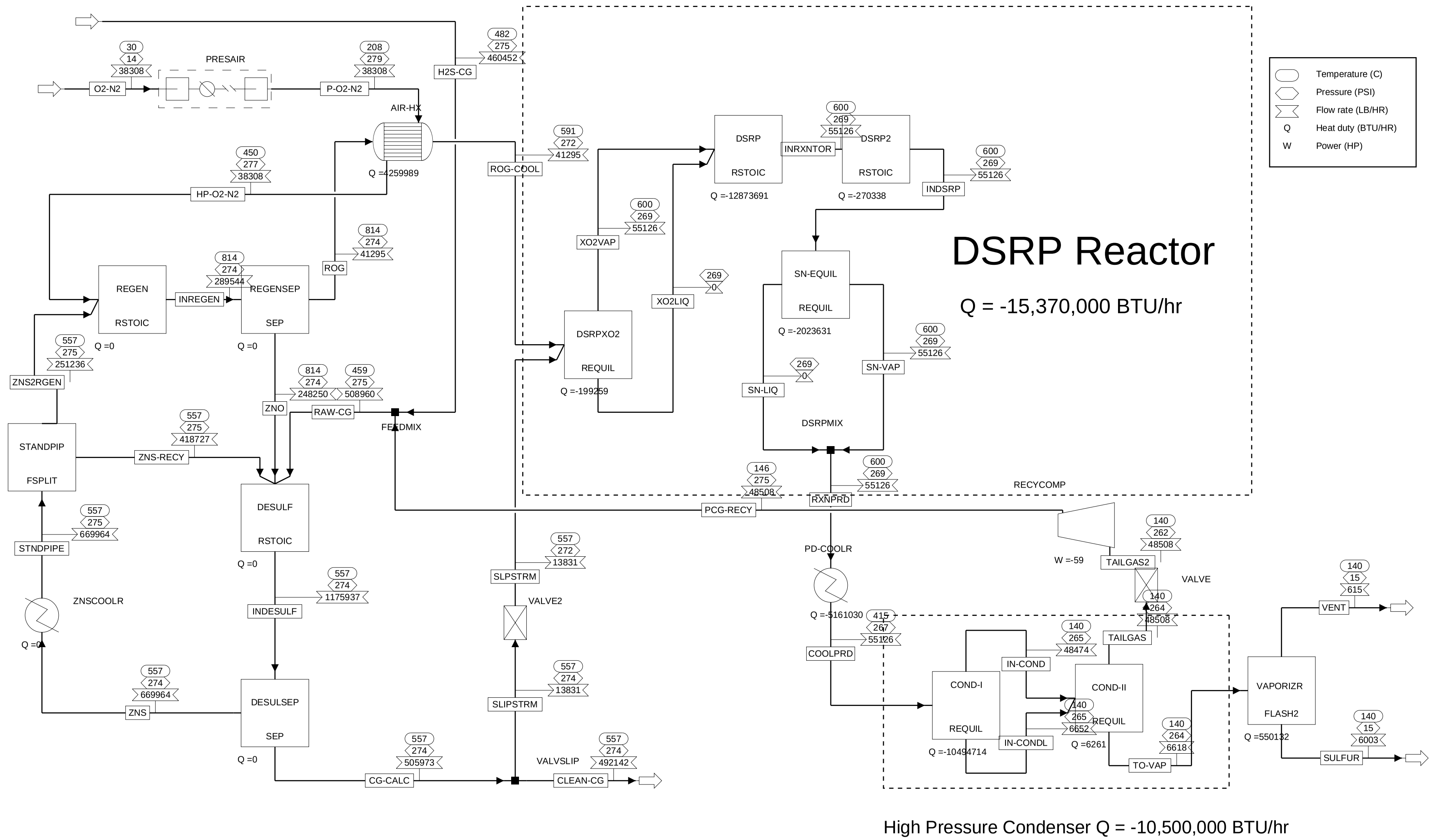
Advanced Hot Gas Process Simulations

AHGP (base case)	0.85 mole% H_2S	260 MW generated
AHGP-b	2.50 mole% H_2S	260 MW generated
AHGP-c	0.25 mole% H_2S	260 MW generated
AHGP-100*	0.85 mole% H_2S	110 MW generated
AHGP-500*	0.85 mole% H_2S	540 MW generated

*DSRP-100, AHGP-100, and AHGP-500 were not simulated. The flowrates and heat duties will scale directly from the base cases (DSRP and AHGP). DSRP-100 and AHGP-100 values equal DSRP and AHGP values scaled by 0.4211. AHGP-500 values equal AHGP values scaled by 2.1055.

DSRP - based Desulfurization

12/20/97 DSRP



DSRP 12/20/97 pg1

Display ALLS	TREAMS	CG-CALC	CLEAN-CG	COOLPRD	H2S-CG	HP-O2-N2	IN-COND	IN-CONDL	INDESULF	INDSRP	INREGEN	INRXNTOR	O2-N2	Display ALLS	TREAMS
Units:	From	DESULSEP	VALVSLIP	PD-COOLR		AIR-HX	COND-I	COND-I	DESULF	DSRP2	REGEN	DSRP		Units:	From
Format: SOLI	DS To	VALVSLIP		COND-I	FEEDMIX	REGEN	COND-II	COND-II	DESULSEP	SN-EQUIL	REGENSEP	DSRP2	PRESAIR	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	557.2	557.2	415	482.2	450	140	140	557.2	600	814.4	600	30	Temperature	[C]
Pressure	[PSI]	274.4	274.4	266.6	275	276.9	264.6	264.6	274.4	268.6	273.6	268.6	13.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	0	0.43	1	0.143	1	1	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0.57	0	0.857	0	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	505972.781	492141.531	55125.902	460451.688	38308.016	48474.25	6651.654	1.18E+06	55125.902	289544.469	55126.063	38308.016	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1.37E+06	1.33E+06	87129.523	1.15E+06	67432.328	49166.316	48.953	1.37E+06	112094.914	96496.891	111308.695	567372.75	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.15E+09	-1.11E+09	-6.57E+07	-1.09E+09	7.31E+06	-7.15E+07	-4.68E+06	-4.84E+09	-5.85E+07	-1.38E+09	-5.82E+07	79241.141	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.37	0.37	0.633	0.399	0.568	0.986	135.877	0.857	0.492	3.001	0.495	0.068	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		63.71	61.968	63.879			63.7	0.18	63.71	63.879	11909.276	1787.733		O2S	
H2S		6.82	6.634	72.073	6270.481		71.958	0.116	6.82	72.072		72.072		H2S	
H2O		78082.43	75947.969	4963.325	70525.25		4175.116	788.209	78082.43	4963.325		4963.325		H2O	
S2		< 0.001	< 0.001	2631.176			< 0.001	0.212	< 0.001	5861.615		3273.283		S2	
S6		0.39	0.38	1891.255			0.965	1890.29	0.39	0.011		0.011		S6	
S8		2.764	2.688	1339.271			2.025	3968.209	2.764	0.076		0.076		S8	
CO		218164.266	212200.516	2.262	218162		2.262	< 0.001	218164.266	2.262		2.262		CO	
CO2		130332.672	126769.898	12929.425	117407.195		12925.243	4.181	130332.672	12929.425		10560.996		CO2	
H2		11766.221	11444.579	0.85	11765.37		0.85	< 0.001	11766.221	0.85		0.85		H2	
O2						8922.588							8922.588	O2	
N2		67553.203	65706.57	31232.063	36321.383	29385.428	31231.807	0.256	67553.203	31232.063	29385.428	31232.063	29385.428	N2	
COS		0.323	0.314	0.323			0.323	< 0.001	0.323	0.323		3233.392		COS	
ZNO											15129.819			ZNO	
ZNS									144457.359		36055			ZNS	
AL2O3									525506.5		197064.938			AL2O3	

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Display ALLS	TREAMS	P-O2-N2	PCG-RECY	RAW-CG	ROG	ROG-COOL	RXNPRD	SLIPSTRM	SLPSTRM	SN-LIQ	SN-VAP	STNDPIPE	SULFUR	Display ALLS	TREAMS
Units:	From	PRESAIR	RECYCOMP	FEEDMIX	REGENSEP	AIR-HX	DSRPMIX	VALVSLIP	VALVE2	SN-EQUIL	SN-EQUIL	ZNSCOOLR	VAPORIZR	Units:	From
Format: SOLI	DS To	AIR-HX	FEEDMIX	DESULF	AIR-HX	DSRPXO2	PD-COOLR	VALVE2	DSRPXO2	DSRPMIX	DSRPMIX	STANDPIP		Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	LIQUID		Phas
Temperature	[C]	207.6	146.5	458.8	814.4	590.7	600	557.2	557.2		600	557.2	140	Temperature	[C]
Pressure	[PSI]	278.9	275	275	273.6	271.6	268.6	274.4	271.6		268.6	275	14.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	1	1		1	0	0	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0		0	1	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	38308.016	48507.805	508959.5	41294.703	41294.703	55125.902	13831.278	13831.278	0	55125.902	669963.875	6002.956	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	44510.953	48127.191	1.20E+06	95300.984	76325.195	109933.391	37429.125	37813.297		109933.273	3189.083	33.124	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	3.05E+06	-7.15E+07	-1.16E+09	-9.58E+06	-1.38E+07	-6.05E+07	-3.13E+07	-31319000	0.00E+00	-6.05E+07	-3.69E+09	-4.57E+05	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.861	1.008	0.423	0.433	0.541	0.501	0.37	0.366		0.501	210.08	181.225	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S			63.71	63.71	11909.276	11909.276	63.879	1.742	1.742		63.879		0.001	O2S	
H2S			71.964	6342.445			72.073	0.186	0.186		72.072		< 0.001	H2S	
H2O			4208.244	74733.492			4963.325	2134.462	2134.462		4963.325		146.193	H2O	
S2			< 0.001	< 0.001			2631.176	trace	trace		2631.176		0.212	S2	
S6			0.39	0.39			1891.255	0.011	0.011		1891.255		724.696	S6	
S8			2.764	2.764			1339.271	0.076	0.076		1339.271		5131.851	S8	
CO			2.262	218164.266			2.262	5963.741	5963.741		2.262		trace	CO	
CO2			12925.476	130332.672			12929.425	3562.775	3562.775		12929.425		0.002	CO2	
H2			0.85	11766.221			0.85	321.642	321.642		0.85		trace	H2	
O2		8922.588												O2	
N2		29385.428	31231.822	67553.203	29385.428	29385.428	31232.063	1846.635	1846.635		31232.063		trace	N2	
COS			0.323	0.323			0.323	0.009	0.009		0.323		trace	COS	
ZNO														ZNO	
ZNS												144457.359		ZNS	
AL2O3												525506.5		AL2O3	

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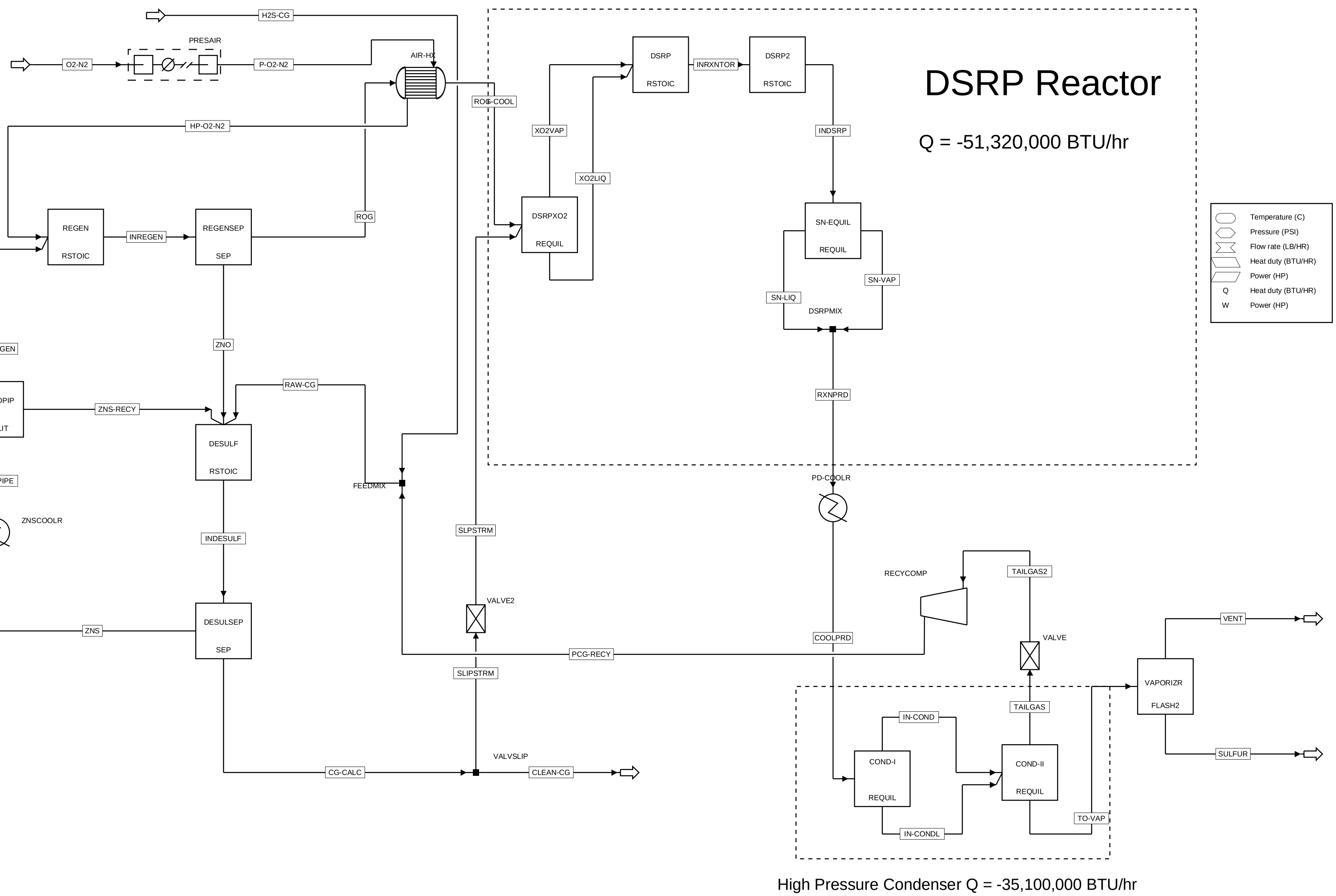
Display ALLS	TREAMS	TAILGAS	TAILGAS2	TO-VAP	VENT	XO2LIQ	XO2VAP	ZNO	ZNS	ZNS-RECY	ZNS2RGEN			Display ALLS	TREAMS
Units:	From	COND-II	VALVE	COND-II	VAPORIZR	DSRPXO2	DSRPXO2	REGENSEP	DESULSEP	STANDPIP	STANDPIP			Units:	From
Format: SOLI	DS To	VALVE	RECYCOMP	VAPORIZR		DSRP	DSRP	DESULF	ZNSCOOLR	DESULF	REGEN			Format: SOLI	DS To
	Phas	VAPOR	VAPOR	LIQUID	VAPOR	MISSING	VAPOR	MISSING	MISSING	MISSING	MISSING				Phas
Temperature	[C]	140	139.9	140	140		600	814.4	557.2	557.2	557.2			Temperature	[C]
Pressure	[PSI]	264.4	262	264.4	14.7		268.6	273.6	274.4	275	275			Pressure	[PSI]
Mass VFrac		1	1	0	1		1	0	0	0	0			Mass VFrac	
Mass SFrac		0	0	0	0		0	1	1	1	1			Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	48507.805	48507.805	6618.099	615.143	0	55126.27	248249.766	669963.875	418727.406	251236.453			Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	49255.441	49703.582	46.323	18288.283		118222.719	1195.906	3189.083	1993.177	1195.906			Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-7.17E+07	-7.17E+07	-4482800	-3.48E+06	0.00E+00	-4.54E+07	-1.37E+09	-3.69E+09	-2307300000	-1.38E+09			Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.985	0.976	142.868	0.034		0.466	207.583	210.08	210.08	210.08			Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		63.71	63.71	0.17	0.169		11911.018							O2S	
H2S		71.964	71.964	0.109	0.109		0.186							H2S	
H2O		4208.244	4208.244	755.081	608.888		1209.115							H2O	
S2		< 0.001	< 0.001	0.212	< 0.001		trace							S2	
S6		0.39	0.39	724.917	0.221		0.011							S6	
S8		2.764	2.764	5133.418	1.567		0.076							S8	
CO		2.262	2.262	< 0.001	< 0.001		4524.921							CO	
CO2		12925.476	12925.476	3.95	3.948		5823.677							CO2	
H2		0.85	0.85	< 0.001	< 0.001		425.192							H2	
O2														O2	
N2		31231.822	31231.822	0.241	0.241		31232.063							N2	
COS		0.323	0.323	< 0.001	< 0.001		0.009							COS	
ZNO								15129.819						ZNO	
ZNS								36055	144457.359	90285.852	54171.508			ZNS	
AL2O3								197064.938	525506.5	328441.563	197064.938			AL2O3	

DSRP-b - based Desulfurization

12/20/97 DSRP-b



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Display ALLS	TREAMS	CG-CALC	CLEAN-CG	COOLPRD	H2S-CG	HP-O2-N2	IN-COND	IN-CONDL	INDESULF	INDSRP	INREGEN	INRXNTOR	O2-N2	Display ALLS	TREAMS
Units:	From	DESULSEP	VALVSLIP	PD-COOLR		AIR-HX	COND-I	COND-I	DESULF	DSRP2	REGEN	DSRP		Units:	From
Format: SOLI	DS To	VALVSLIP		COND-I	FEEDMIX	REGEN	COND-II	COND-II	DESULSEP	SN-EQUIL	REGENSEP	DSRP2	PRESAIR	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	647.5	647.5	415	482.2	450	140	140	647.5	600	810	600	30	Temperature	[C]
Pressure	[PSI]	273.6	273.6	265.6	275	277.4	263.6	263.6	273.6	267.6	272.6	267.6	13.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	0	0.197	1	0.099	1	1	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0.803	0	0.901	0	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	655189.75	601554.75	184610.219	501234.031	121500	163323.219	21287.002	3324180	184607.219	1322550	184607.719	121500	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1887870	1733330	294846.875	1244460	213489.484	166528.75	159.053	1900580	378670.75	307882.375	376264.813	1799510	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.36E+09	-1.24E+09	-2.24E+08	-1.16E+09	2.32E+07	-2.43E+08	-1.61E+07	-1.60E+10	-2.00E+08	-6.55E+09	-1.99E+08	2.51E+05	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.347	0.347	0.626	0.403	0.569	0.981	133.836	1.749	0.488	4.296	0.491	0.068	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		194.606	178.675	195.105			194.579	0.526	194.606	195.105	37772.176	5444.741		O2S	
H2S		62.453	57.341	240.342	19916.91		239.971	0.371	62.453	240.342		240.342		H2S	
H2O		99935.453	91754.563	17155.324	74738.914		14470.859	2684.464	99935.453	17155.324		17155.324		H2O	
S2		< 0.001	< 0.001	8658.119			< 0.001	0.695	< 0.001	18593.258		10711.036		S2	
S6		1.28	1.175	5885.709			3.086	5882.623	1.28	0.105		0.105		S6	
S8		8.967	8.233	4053.262			6.665	12704.021	8.967	0.734		0.734		S8	
CO		231203.391	212276.672	6.89	231196.5		6.89	< 0.001	231203.391	6.89		6.89		CO	
CO2		167878.578	154135.734	43469.434	124421.914		43455.953	13.482	167878.578	43469.434		36256.879		CO2	
H2		12471.098	11450.191	2.783	12468.315		2.783	< 0.001	12471.098	2.783		2.783		H2	
O2						28299.416							28299.416	O2	
N2		143432.969	131691.281	104942.266	38491.473	93200.586	104941.445	0.818	143432.969	104942.266	93200.586	104942.266	93200.586	N2	
COS		0.984	0.904	0.985			0.984	< 0.001	0.984	0.985		9846.616		COS	
ZNO											47986.641			ZNO	
ZNS									576614.875		202017.281			ZNS	
AL2O3									2092380		941570.188			AL2O3	

DSRP-b 12/20/97 pg.2

Display ALLS	TREAMS	P-O2-N2	PCG-RECY	RAW-CG	ROG	ROG-COOL	RXNPRD	SLIPSTRM	SLPSTRM	SN-LIQ	SN-VAP	STNDPIPE	SULFUR	Display ALLS	TREAMS
Units:	From	PRESAIR	RECYCOMP	FEEDMIX	REGENSEP	AIR-HX	DSRPMIX	VALVSLIP	VALVE2	SN-EQUIL	SN-EQUIL	ZNSCOOLR	VAPORIZR	Units:	From
Format: SOLI	DS To	AIR-HX	FEEDMIX	DESULF	AIR-HX	DSRPXO2	PD-COOLR	VALVE2	DSRPXO2	DSRPMIX	DSRPMIX	STANDPIP		Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	LIQUID		Phas
Temperature	[C]	207.7	147.5	418.7	810	586.3	600	647.5	647.5		600	640	140	Temperature	[C]
Pressure	[PSI]	279.4	275	275	272.6	270.6	267.6	273.6	270.6		267.6	275	14.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	1	1		1	0	0	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0		0	1	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	121500	163428.5	664662.5	130972.758	130972.758	184610.219	53635.004	53635.004	0	184610.219	2668990	19044.756	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	140938.547	162790.359	1410320	302165.188	241725.797	372001.438	154544.391	156250.734		372001.031	12704.87	105.13	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	9691570	-2.43E+08	-1.40E+09	-3.07E+07	-4.42E+07	-2.06E+08	-1.11E+08	-1.11E+08	0	-2.06E+08	-1.46E+10	-1.45E+06	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.862	1.004	0.471	0.433	0.542	0.496	0.347	0.343		0.496	210.076	181.154	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S			194.606	194.606	37772.176	37772.176	195.105	15.931	15.931		195.105		0.002	O2S	
H2S			239.99	20156.9			240.342	5.113	5.113		240.342		0.001	H2S	
H2O			14574.853	89313.766			17155.324	8180.895	8180.895		17155.324		464.116	H2O	
S2			< 0.001	< 0.001			8658.119	< 0.001	< 0.001		8658.119		0.695	S2	
S6			1.28	1.28			5885.709	0.105	0.105		5885.709		2321.039	S6	
S8			8.967	8.967			4053.262	0.734	0.734		4053.262		16258.897	S8	
CO			6.89	231203.391			6.89	18926.723	18926.723		6.89		trace	CO	
CO2			43456.652	167878.578			43469.434	13742.839	13742.839		43469.434		0.006	CO2	
H2			2.783	12471.098			2.783	1020.906	1020.906		2.783		trace	H2	
O2		28299.416												O2	
N2		93200.586	104941.492	143432.969	93200.586	93200.586	104942.266	11741.678	11741.678		104942.266		< 0.001	N2	
COS			0.984	0.984			0.985	0.081	0.081		0.985		trace	COS	
ZNO														ZNO	
ZNS												576614.875		ZNS	
AL2O3												2092380		AL2O3	

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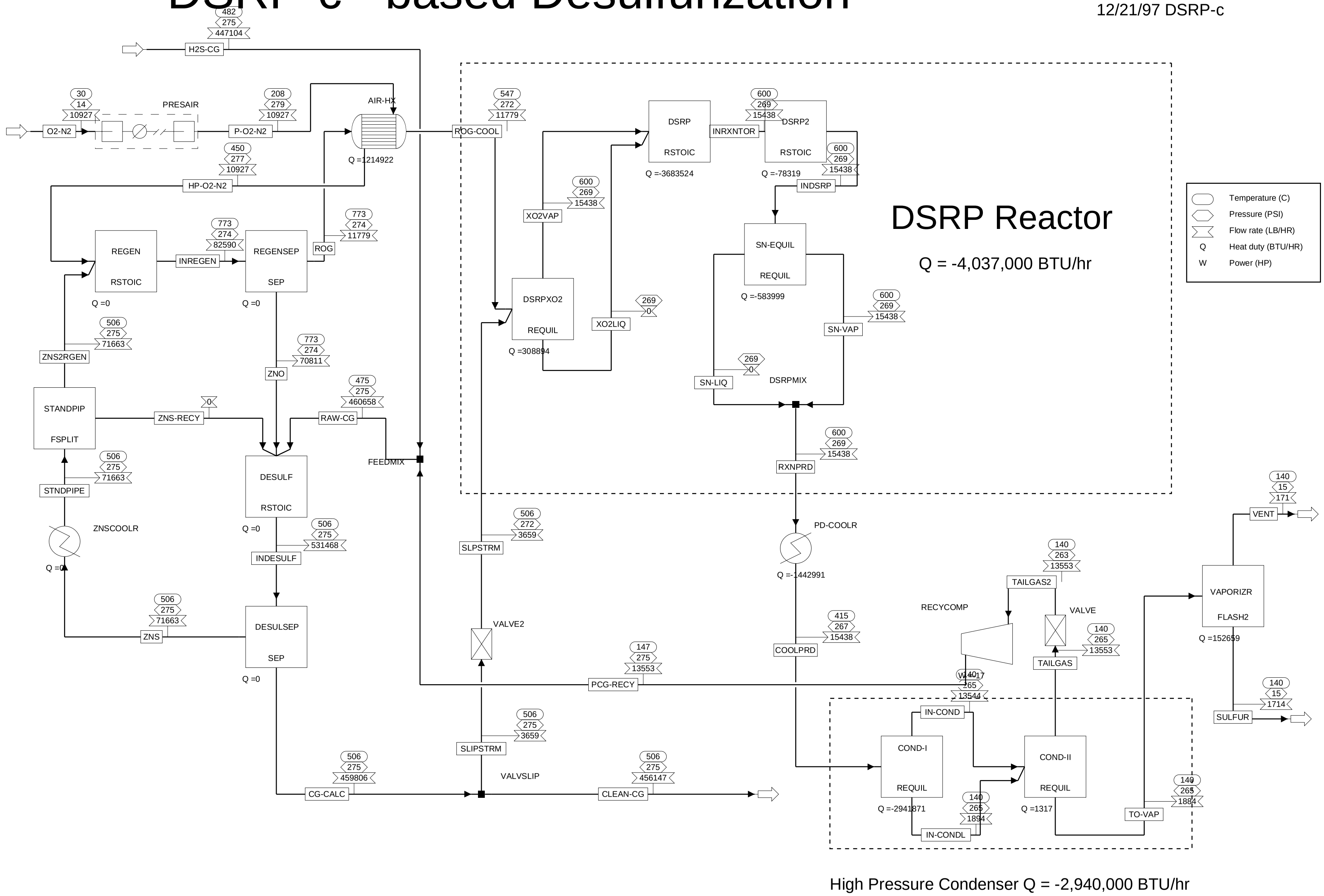
Display ALLS	TREAMS	TAILGAS	TAILGAS2	TO-VAP	VENT	XO2LIQ	XO2VAP	ZNO	ZNS	ZNS-RECY	ZNS2RGEN			Display ALLS	TREAMS
Units:	From	COND-II	VALVE	COND-II	VAPORIZR	DSRPXO2	DSRPXO2	REGENSEP	DESULSEP	STANDPIP	STANDPIP			Units:	From
Format: SOLI	DS To	VALVE	RECYCOMP	VAPORIZR		DSRP	DSRP	DESULF	ZNSCOOLR	DESULF	REGEN			Format: SOLI	DS To
	Phas	VAPOR	VAPOR	LIQUID	VAPOR	MISSING	VAPOR	MISSING	MISSING	MISSING	MISSING				Phas
Temperature	[C]	140	139.9	140	140		600	810	647.5	640	640			Temperature	[C]
Pressure	[PSI]	263.6	261	263.6	14.7		267.6	272.6	273.6	275	275			Pressure	[PSI]
Mass VFrac		1	1	0	1		1	0	0	0	0			Mass VFrac	
Mass SFrac		0	0	0	0		0	1	1	1	1			Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	163428.5	163428.5	21181.719	2136.962	0	184608.375	1191570	2668990	1467950	1201050			Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	166691.297	168340.75	150.956	63551.742		398196.594	5717.191	12704.87	6987.678	5717.191			Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-2.43E+08	-2.43E+08	-1.54E+07	-1.21E+07	0	-1.59E+08	-6.52E+09	-1.46E+10	-8.03E+09	-6.57E+09			Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.98	0.971	140.317	0.034		0.464	208.419	210.076	210.076	210.076			Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		194.606	194.606	0.499	0.497		37788.105							O2S	
H2S		239.99	239.99	0.353	0.352		5.113							H2S	
H2O		14574.853	14574.853	2580.471	2116.355		4870.572							H2O	
S2		< 0.001	< 0.001	0.695	< 0.001		< 0.001							S2	
S6		1.28	1.28	2321.814	0.776		0.105							S6	
S8		8.967	8.967	16264.332	5.434		0.734							S8	
CO		6.89	6.89	< 0.001	< 0.001		13779.622							CO	
CO2		43456.652	43456.652	12.781	12.774		21830.447							CO2	
H2		2.783	2.783	< 0.001	< 0.001		1391.339							H2	
O2														O2	
N2		104941.492	104941.492	0.773	0.773		104942.266							N2	
COS		0.984	0.984	< 0.001	< 0.001		0.081							COS	
ZNO								47986.641						ZNO	
ZNS								202017.281	576614.875	317138.188	259476.688			ZNS	
AL2O3								941570.188	2092380	1150810	941570.188			AL2O3	

DSRP-c - based Desulfurization

12/21/97 DSRP-c



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Display ALLS	TREAMS	CG-CALC	CLEAN-CG	COOLPRD	H2S-CG	HP-O2-N2	IN-COND	IN-CONDL	INDESULF	INDSRP	INREGEN	INRXNTOR	O2-N2	Display ALLS	TREAMS
Units:	From	DESULSEP	VALVSLIP	PD-COOLR		AIR-HX	COND-I	COND-I	DESULF	DSRP2	REGEN	DSRP		Units:	From
Format: SOLI	DS To	VALVSLIP		COND-I	FEEDMIX	REGEN	COND-II	COND-II	DESULSEP	SN-EQUIL	REGENSEP	DSRP2	PRESAIR	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	506.4	506.4	415	482.2	450	140	140	506.4	600	772.6	600	30	Temperature	[C]
Pressure	[PSI]	274.9	274.9	267.1	275	277.3	265.1	265.1	274.9	269.1	274.1	269.1	13.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	0	0.865	1	0.143	1	1	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0.135	0	0.857	0	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	459805.594	456146.75	15437.757	447104.125	10927	13544.205	1893.551	531468.188	15437.757	82589.578	15437.803	10927	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1187320	1177870	24296.518	1125490	19206.873	13705.025	13.869	1187660	31278.654	26436.727	31051.443	161837.719	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.08E+09	-1.08E+09	-1.83E+07	-1.07E+09	2.09E+06	-1.99E+07	-1.30E+06	-1.48E+09	-1.63E+07	-3.94E+08	-1.62E+07	2.26E+04	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.387	0.387	0.635	0.397	0.569	0.988	136.533	0.447	0.494	3.124	0.497	0.068	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		15.704	15.579	15.747			15.702	0.045	15.704	15.747	3397.009	515.087		O2S	
H2S		14.867	14.749	20.399	1801.676		20.366	0.033	14.867	20.399		20.399		H2S	
H2O		71265.617	70698.523	1374.565	69146.555		1154.749	219.816	71265.617	1374.565		1374.565		H2O	
S2		< 0.001	< 0.001	741.13			< 0.001	0.06	< 0.001	1673.231		923.483		S2	
S6		0.11	0.109	543.432			0.274	543.158	0.11	0.001		0.001		S6	
S8		0.78	0.774	388.677			0.57	1129.177	0.78	0.006		0.006		S8	
CO		213897.828	212195.766	0.655	213897.172		0.655	trace	213897.828	0.655		0.655		CO	
CO2		118728.914	117784.148	3618.024	115112.016		3616.834	1.189	118728.914	3618.024		2931.973		CO2	
H2		11535.61	11443.817	0.24	11535.37		0.24	trace	11535.61	0.24		0.24		H2	
O2						2545.084							2545.084	O2	
N2		44346.063	43993.184	8734.794	35611.336	8381.916	8734.721	0.073	44346.063	8734.794	8381.916	8734.794	8381.916	N2	
COS		0.094	0.093	0.094			0.094	< 0.001	0.094	0.094		936.6		COS	
ZNO											4315.638			ZNO	
ZNS									15452.001		10284.438			ZNS	
AL2O3									56210.578		56210.578			AL2O3	

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Display ALLS	TREAMS	P-O2-N2	PCG-RECY	RAW-CG	ROG	ROG-COOL	RXNPRD	SLIPSTRM	SLPSTRM	SN-LIQ	SN-VAP	STNDPIPE	SULFUR	Display ALLS	TREAMS
Units:	From	PRESAIR	RECYCOMP	FEEDMIX	REGENSEP	AIR-HX	DSRPMIX	VALVSLIP	VALVE2	SN-EQUIL	SN-EQUIL	ZNSCOOLR	VAPORIZR	Units:	From
Format: SOLI	DS To	AIR-HX	FEEDMIX	DESULF	AIR-HX	DSRPXO2	PD-COOLR	VALVE2	DSRPXO2	DSRPMIX	DSRPMIX	STANDPIP		Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	LIQUID		Phas
Temperature	[C]	207.7	146.8	475.2	772.6	547.5	600	506.4	506.4		600	506.4	140	Temperature	[C]
Pressure	[PSI]	279.3	275	275	274.1	272.1	269.1	274.9	272.1		269.1	275	14.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	1	1		1	0	0	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0		0	1	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	10927	13553.382	460657.5	11778.925	11778.925	15437.757	3658.844	3658.844	0	15437.757	71662.578	1713.577	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	12679.411	13452.904	1139190	26095.605	20645.352	30656.066	9447.959	9544.736		30656.035	341.12	9.454	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	871552.063	-1.99E+07	-1.09E+09	-2.96E+06	-4.18E+06	-1.68E+07	-8.63E+06	-8.63E+06	0	-1.68E+07	-3.97E+08	-1.30E+05	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.862	1.007	0.404	0.451	0.571	0.504	0.387	0.383		0.504	210.08	181.246	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S			15.704	15.704	3397.009	3397.009	15.747	0.125	0.125		15.747		< 0.001	O2S	
H2S			20.368	1822.044			20.399	0.118	0.118		20.399		< 0.001	H2S	
H2O			1163.806	70310.359			1374.565	567.087	567.087		1374.565		41.723	H2O	
S2			< 0.001	< 0.001			741.13	trace	trace		741.13		0.06	S2	
S6			0.11	0.11			543.432	0.001	0.001		543.432		206.281	S6	
S8			0.78	0.78			388.677	0.006	0.006		388.677		1465.512	S8	
CO			0.655	213897.828			0.655	1702.064	1702.064		0.655			CO	
CO2			3616.9	118728.914			3618.024	944.77	944.77		3618.024		0.001	CO2	
H2			0.24	11535.61			0.24	91.793	91.793		0.24		trace	H2	
O2		2545.084												O2	
N2		8381.916	8734.726	44346.063	8381.916	8381.916	8734.794	352.878	352.878		8734.794		trace	N2	
COS			0.094	0.094			0.094	0.001	0.001		0.094		trace	COS	
ZNO														ZNO	
ZNS												15452.001		ZNS	
AL2O3												56210.578		AL2O3	

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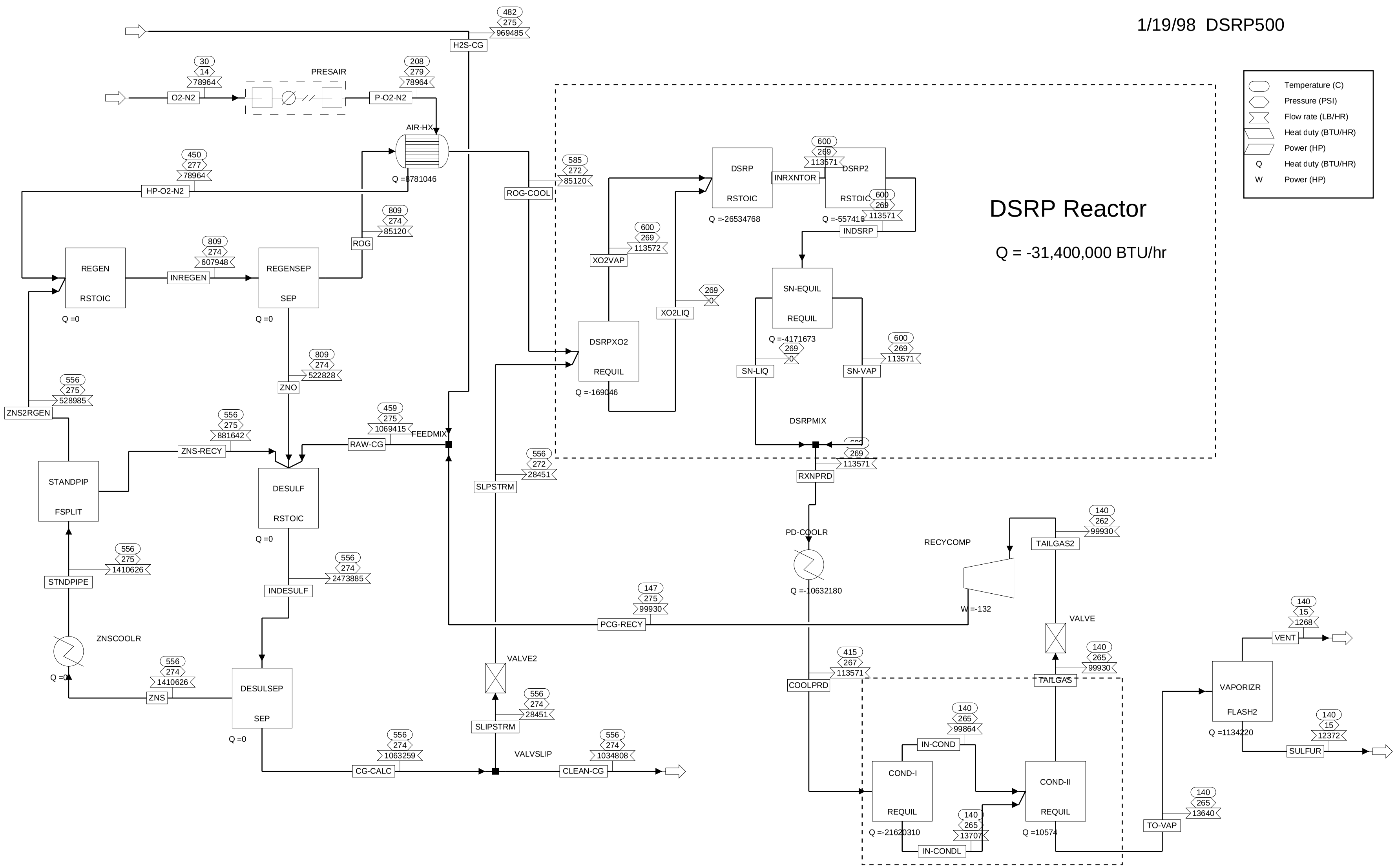
Display ALLS	TREAMS	TAILGAS	TAILGAS2	TO-VAP	VENT	XO2LIQ	XO2VAP	ZNO	ZNS	ZNS-RECY	ZNS2RGEN			Display ALLS	TREAMS
Units:	From	COND-II	VALVE	COND-II	VAPORIZR	DSRPXO2	DSRPXO2	REGENSEP	DESULSEP	STANDPIP	STANDPIP			Units:	From
Format: SOLI	DS To	VALVE	RECYCOMP	VAPORIZR		DSRP	DSRP	DESULF	ZNSCOOLR	DESULF	REGEN			Format: SOLI	DS To
	Phas	VAPOR	VAPOR	LIQUID	VAPOR	MISSING	VAPOR	MISSING	MISSING	MISSING	MISSING				Phas
Temperature	[C]	140	139.9	140	140		600	772.6	506.4		506.4			Temperature	[C]
Pressure	[PSI]	265.1	262.5	265.1	14.7		269.1	274.1	274.9		275			Pressure	[PSI]
Mass VFrac		1	1	0	1		1	0	0		0			Mass VFrac	
Mass SFrac		0	0	0	0		0	1	1		1			Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	13553.382	13553.382	1884.374	170.798	0	15437.86	70810.656	71662.578	0	71662.578			Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	13719.107	13854.065	13.12	5077.429		33024.156	341.12	341.12		341.12			Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-2.00E+07	-2.00E+07	-1.25E+06	-9.65E+05	0	-1.25E+07	-3.91E+08	-3.97E+08	0.00E+00	-3.97E+08			Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.988	0.978	143.629	0.034		0.467	207.583	210.08		210.08			Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		15.704	15.704	0.043	0.042		3397.134							O2S	
H2S		20.368	20.368	0.031	0.031		0.118							H2S	
H2O		1163.806	1163.806	210.759	169.036		315.4							H2O	
S2		< 0.001	< 0.001	0.06	< 0.001		trace							S2	
S6		0.11	0.11	206.342	0.061		0.001							S6	
S8		0.78	0.78	1465.947	0.435		0.006							S8	
CO		0.655	0.655	trace	trace		1310.713							CO	
CO2		3616.9	3616.9	1.124	1.123		1559.735							CO2	
H2		0.24	0.24	trace	trace		119.958							H2	
O2														O2	
N2		8734.726	8734.726	0.069	0.069		8734.794							N2	
COS		0.094	0.094	< 0.001	< 0.001		0.001							COS	
ZNO								4315.638						ZNO	
ZNS								10284.438	15452.001		15452.001			ZNS	
AL2O3								56210.578	56210.578		56210.578			AL2O3	

DSRP-500 - based Desulfurization

1/19/98 DSRP500



High Pressure Condenser Q = -21,600,000 BTU/hr

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Display ALLS	TREAMS	CG-CALC	CLEAN-CG	COOLPRD	H2S-CG	HP-O2-N2	IN-COND	IN-CONDL	INDESULF	INDSRP	INREGEN	INRXNTOR	O2-N2	Display ALLS	TREAMS
Units:	From	DESULSEP	VALVSLIP	PD-COOLR		AIR-HX	COND-I	COND-I	DESULF	DSRP2	REGEN	DSRP		Units:	From
Format: SOLI	DS To	VALVSLIP		COND-I	FEEDMIX	REGEN	COND-II	COND-II	DESULSEP	SN-EQUIL	REGENSEP	DSRP2	PRESAIR	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	556.1	556.1	415	482.2	450	140	140	556.1	600	808.9	600	30	Temperature	[C]
Pressure	[PSI]	274.4	274.4	266.6	275	276.9	264.6	264.6	274.4	268.6	273.6	268.6	13.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	0	0.43	1	0.14	1	1	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0.57	0	0.86	0	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	1063260	1034810	113570.906	969485	78963.703	99863.633	13707.267	2.47E+06	113570.898	607948.188	113571.234	78963.703	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	2.87E+06	2.80E+06	179490.672	2.43E+06	138997.172	101285.656	100.854	2.88E+06	230923.609	197974.375	229302.5	1169520	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-2.41E+09	-2.35E+09	-1.35E+08	-2.30E+09	1.51E+07	-1.47E+08	-9.63E+06	-1.02E+10	-1.20E+08	-2.90E+09	-1.20E+08	163338.5	Entha lpy	[BTU/HR]
Density	[LB/CUFT]	0.37	0.37	0.633	0.399	0.568	0.986	135.912	0.859	0.492	3.071	0.495	0.068	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S		134.563	130.963	134.922			134.543	0.379	134.563	134.922	24548.4	3689.37		O2S	
H2S		298.899	290.901	156.105	13202.553		155.855	0.25	298.899	156.105		156.105		H2S	
H2O		164058.219	159668.297	10220.242	148491.516		8597.656	1622.586	164058.219	10220.242		10220.242		H2O	
S2		< 0.001	< 0.001	5421.442			< 0.001	0.438	< 0.001	12080.898		6743.967		S2	
S6		0.803	0.782	3898.438			1.989	3896.449	0.803	0.021		0.021		S6	
S8		5.689	5.537	2761.193			4.175	8178.022	5.689	0.152		0.152		S8	
CO		459346.656	447055.344	4.665	459341.969		4.665	< 0.001	459346.656	4.665		4.665		CO	
CO2		273825.563	266498.469	26631.824	247201.875		26623.211	8.614	273825.563	26631.824		21748.314		CO2	
H2		24773.842	24110.938	1.752	24772.09		1.752	< 0.001	24773.842	1.752		1.752		H2	
O2						18391.988							18391.988	O2	
N2		140814.141	137046.203	64339.652	76474.984	60571.711	64339.125	0.528	140814.141	64339.652	60571.711	64339.652	60571.711	N2	
COS		0.666	0.649	0.667			0.666	< 0.001	0.666	0.667		6666.993		COS	
ZNO											31186.855			ZNO	
ZNS									304172		76721.18			ZNS	
AL2O3									1106450		414920.063			AL2O3	

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Display ALLS	TREAMS	P-O2-N2	PCG-RECY	RAW-CG	ROG	ROG-COOL	RXNPRD	SLIPSTRM	SLPSTRM	SN-LIQ	SN-VAP	STNDPIPE	SULFUR	Display ALLS	TREAMS
Units:	From	PRESAIR	RECYCOMP	FEEDMIX	REGENSEP	AIR-HX	DSRPMIX	VALVSLIP	VALVE2	SN-EQUIL	SN-EQUIL	ZNSCOOLR	VAPORIZR	Units:	From
Format: SOLI	DS To	AIR-HX	FEEDMIX	DESULF	AIR-HX	DSRPXO2	PD-COOLR	VALVE2	DSRPXO2	DSRPMIX	DSRPMIX	STANDPIP		Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	LIQUID		Phas
Temperature	[C]	207.6	147.1	459.3	808.9	585.1	600	556.1	556.1		600	556.1	140	Temperature	[C]
Pressure	[PSI]	278.9	275	275	273.6	271.6	268.6	274.4	271.6		268.6	275	14.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	1	1		1	0	0	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0		0	1	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	78963.703	99930.422	1069420	85120.109	85120.109	113570.906	28450.951	28450.951	0	113570.906	1410630	12372.203	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	91749.711	99302.75	2.53E+06	195456.375	156297.594	226467.672	76921.539	77711.063		226467.438	6714.697	68.27	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	6.30E+06	-1.47E+08	-2.44E+09	-2.00E+07	-2.88E+07	-1.25E+08	-6.45E+07	-64473000	0.00E+00	-1.25E+08	-7.77E+09	-9.42E+05	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.861	1.006	0.422	0.435	0.545	0.501	0.37	0.366		0.501	210.08	181.225	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S			134.563	134.563	24548.4	24548.4	134.922	3.601	3.601		134.922		0.002	O2S	
H2S			155.868	13358.421			156.105	7.998	7.998		156.105		0.001	H2S	
H2O			8663.586	157155.109			10220.242	4389.911	4389.911		10220.242		301.303	H2O	
S2			< 0.001	< 0.001			5421.442	trace	trace		5421.442		0.437	S2	
S6			0.803	0.803			3898.438	0.021	0.021		3898.438		1493.655	S6	
S8			5.689	5.689			2761.193	0.152	0.152		2761.193		10576.802	S8	
CO			4.665	459346.656			4.665	12291.313	12291.313		4.665		trace	CO	
CO2			26623.678	273825.563			26631.824	7327.093	7327.093		26631.824		0.005	CO2	
H2			1.752	24773.842			1.752	662.905	662.905		1.752		trace	H2	
O2		18391.988												O2	
N2		60571.711	64339.152	140814.141	60571.711	60571.711	64339.652	3767.94	3767.94		64339.652		trace	N2	
COS			0.666	0.666			0.667	0.018	0.018		0.667		trace	COS	
ZNO														ZNO	
ZNS												304172		ZNS	
AL2O3												1106450		AL2O3	

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Display ALLS	TREAMS	P-O2-N2	PCG-RECY	RAW-CG	ROG	ROG-COOL	RXNPRD	SLIPSTRM	SLPSTRM	SN-LIQ	SN-VAP	STNDPIPE	SULFUR	Display ALLS	TREAMS
Units:	From	PRESAIR	RECYCOMP	FEEDMIX	REGENSEP	AIR-HX	DSRPMIX	VALVSLIP	VALVE2	SN-EQUIL	SN-EQUIL	ZNSCOOLR	VAPORIZR	Units:	From
Format: SOLI	DS To	AIR-HX	FEEDMIX	DESULF	AIR-HX	DSRPXO2	PD-COOLR	VALVE2	DSRPXO2	DSRPMIX	DSRPMIX	STANDPIP		Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	LIQUID		Phas
Temperature	[C]	207.6	147.1	459.3	808.9	585.1	600	556.1	556.1		600	556.1	140	Temperature	[C]
Pressure	[PSI]	278.9	275	275	273.6	271.6	268.6	274.4	271.6		268.6	275	14.7	Pressure	[PSI]
Mass VFrac		1	1	1	1	1	1	1	1		1	0	0	Mass VFrac	
Mass SFrac		0	0	0	0	0	0	0	0		0	1	0	Mass SFrac	
*** ALL PHAS	ES ***													*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	78963.703	99930.422	1069420	85120.109	85120.109	113570.906	28450.951	28450.951	0	113570.906	1410630	12372.203	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	91749.711	99302.75	2.53E+06	195456.375	156297.594	226467.672	76921.539	77711.063		226467.438	6714.697	68.27	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	6.30E+06	-1.47E+08	-2.44E+09	-2.00E+07	-2.88E+07	-1.25E+08	-6.45E+07	-64473000	0.00E+00	-1.25E+08	-7.77E+09	-9.42E+05	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.861	1.006	0.422	0.435	0.545	0.501	0.37	0.366		0.501	210.08	181.225	Density	[LB/CUFT]
Mass Flow	[LB/HR]													Mass Flow	[LB/HR]
O2S			134.563	134.563	24548.4	24548.4	134.922	3.601	3.601		134.922		0.002	O2S	
H2S			155.868	13358.421			156.105	7.998	7.998		156.105		0.001	H2S	
H2O			8663.586	157155.109			10220.242	4389.911	4389.911		10220.242		301.303	H2O	
S2			< 0.001	< 0.001			5421.442	trace	trace		5421.442		0.437	S2	
S6			0.803	0.803			3898.438	0.021	0.021		3898.438		1493.655	S6	
S8			5.689	5.689			2761.193	0.152	0.152		2761.193		10576.802	S8	
CO			4.665	459346.656			4.665	12291.313	12291.313		4.665		trace	CO	
CO2			26623.678	273825.563			26631.824	7327.093	7327.093		26631.824		0.005	CO2	
H2			1.752	24773.842			1.752	662.905	662.905		1.752		trace	H2	
O2		18391.988												O2	
N2		60571.711	64339.152	140814.141	60571.711	60571.711	64339.652	3767.94	3767.94		64339.652		trace	N2	
COS			0.666	0.666			0.667	0.018	0.018		0.667		trace	COS	
ZNO														ZNO	
ZNS												304172		ZNS	
AL2O3												1106450		AL2O3	

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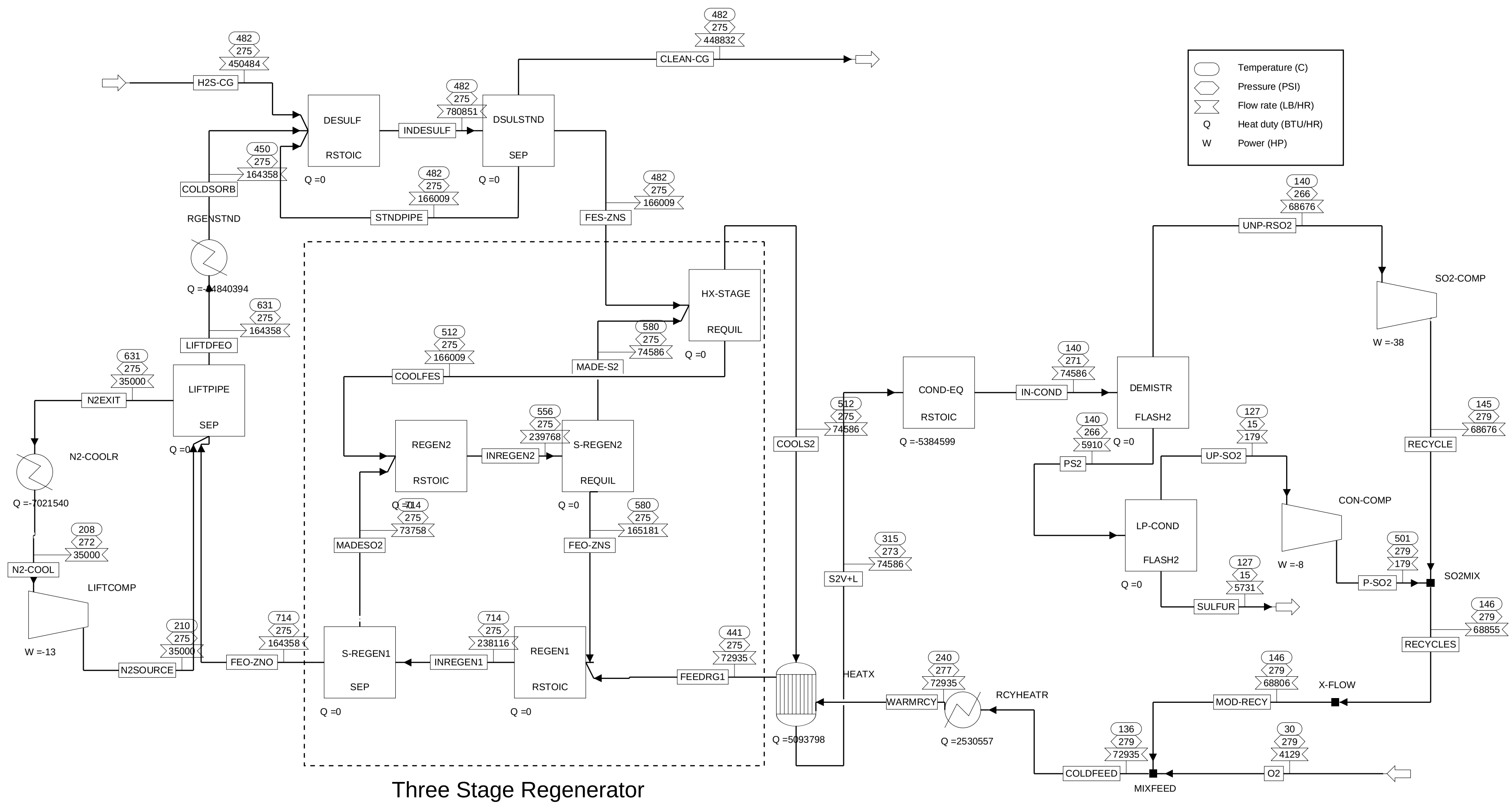

Three Stage Regenerator

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Display ALLS	TREAMS	CLEAN-CG	COLDFEED	COLDSORB	COOLFES	COOLS2	FEEDRG1	FEO-ZNO	FEO-ZNS	FES-ZNS	H2S-CG	IN-COND	Display ALLS	TREAMS
Units:	From	DSULSTND	MIXFEED	RGENSTND	HX-STAGE	HX-STAGE	HEATX	S-REGEN1	S-REGEN2	DSULSTND		COND-EQ	Units:	From
Format: SOLI	DS To		RCYHEATR	DESULF	REGEN2	HEATX	REGEN1	LIFTPIPE	REGEN1	HX-STAGE	DESULF	DEMISTR	Format: SOLI	DS To
	Phas	VAPOR	MIXED	MISSING	MISSING	VAPOR	VAPOR	MISSING	MISSING	MISSING	VAPOR	MIXED		Phas
Temperature	[C]	482.4	136.5	450	512.3	512.3	440.6	713.9	580.4	482.4	482.2	140	Temperature	[C]
Pressure	[PSI]	274.7	279.2	275	274.7	274.7	275.2	274.7	274.7	274.7	275	270.7	Pressure	[PSI]
Mass VFrac		1	1	0	0	1	1	0	0	0	1	0.921	Mass VFrac	
Mass SFrac		< 0.001	0	1	1	0	0	1	1	1	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	448832.344	72935.094	164357.922	166009.453	74586.227	72935.094	164357.922	165181.094	166009.453	450483.875	74586.227	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1.13E+06	30858.555	756.208	793.141	60581.492	59761.762	756.208	768.468	793.141	1.13E+06	2.82E+04	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.09E+09	-1.35E+08	-1.06E+09	-1.03E+09	-1.25E+08	-1.28E+08	-1.04E+09	-1.04E+09	-1.04E+09	-1.07E+09	-1.35E+08	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.397	2.364	217.345	209.306	1.231	1.22	217.345	214.949	209.306	0.399	2.645	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			68800			68860.266	68800					68860.266	O2S	
H2S		23.01									6134.738		H2S	
H2O		73613.648									68998.523		H2O	
S2						519.707							S2	
S6			2.397			2171.238	2.397					2181.632	S6	
S8			3.894			3035.012	3.894					3544.325	S8	
CO		213439.25									213439.25		CO	
CO2		114865.578									114865.578		CO2	
H2		11355.747									11510.675		H2	
O2			4128.805				4128.805						O2	
N2		35535.098									35535.098		N2	
COS													COS	
ZNO				2084.988				2084.988					ZNO	
ZNS					2496.573				2496.573	2496.573			ZNS	
FE2O3				12272.938				12272.938	8199.015				FE2O3	
FEO													FEO	
FES		0.001			13512.88				4485.513	13512.88			FES	
AL2O3				150000	150000			150000	150000	150000			AL2O3	

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Display ALLS	TREAMS	INDESULF	INREGEN1	INREGEN2	LIFTDFEO	MADE-S2	MADESO2	MOD-RECY	N2-COOL	N2EXIT	N2SOURCE	O2	Display ALLS	TREAMS
Units:	From	DESULF	REGEN1	REGEN2	LIFTPIPE	S-REGEN2	S-REGEN1	X-FLOW	N2-COOLR	LIFTPIPE	LIFTCOMP		Units:	From
Format: SOLI	DS To	DSULSTND	S-REGEN1	S-REGEN2	RGENSTND	HX-STAGE	REGEN2	MIXFEED	LIFTCOMP	N2-COOLR	LIFTPIPE	MIXFEED	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	482.4	713.9	556.4	631	580.4	713.9	146.1	208	631	210.1	30	Temperature	[C]
Pressure	[PSI]	274.7	274.7	274.7	274.7	274.7	274.7	279.2	272	274.7	275	279.2	Pressure	[PSI]
Mass VFrac		0.575	0.31	0.311	0	1	1	1	1	1	1	1	Mass VFrac	
Mass SFrac		0.425	0.69	0.689	1	0	0	0	0	0	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	780850.875	238116.203	239767.375	164357.922	74586.227	73758.273	68806.289	35000	35000	35000	4128.805	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1132710	8.08E+04	68028.672	756.208	66784.258	80083.75	27821.109	43008.844	79902.289	42726.648	2669.199	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-3.16E+09	-1.17E+09	-1.16E+09	-1.05E+09	-1.22E+08	-1.30E+08	-1.35E+08	2866710	9.89E+06	2899740	-1712.78	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.689	2.946	3.525	217.345	1.117	0.921	2.473	0.814	0.438	0.819	1.547	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			73722.469	68860.266		68860.266	73722.469	68800					O2S	
H2S		23.01											H2S	
H2O		73613.648											H2O	
S2				5726		1563.893							S2	
S6						2146.613		2.397					S6	
S8						2015.451		3.894					S8	
CO		213439.25											CO	
CO2		114865.578											CO2	
H2		11355.747											H2	
O2			35.801				35.801					4128.805	O2	
N2		35535.098							35000	35000	35000		N2	
COS													COS	
ZNO			2084.988		2084.988								ZNO	
ZNS		4993.106		2496.573									ZNS	
FE2O3			12272.938	8199.016	12272.938								FE2O3	
FEO													FEO	
FES		27025.441		4485.513									FES	
AL2O3		300000	150000	150000	150000								AL2O3	

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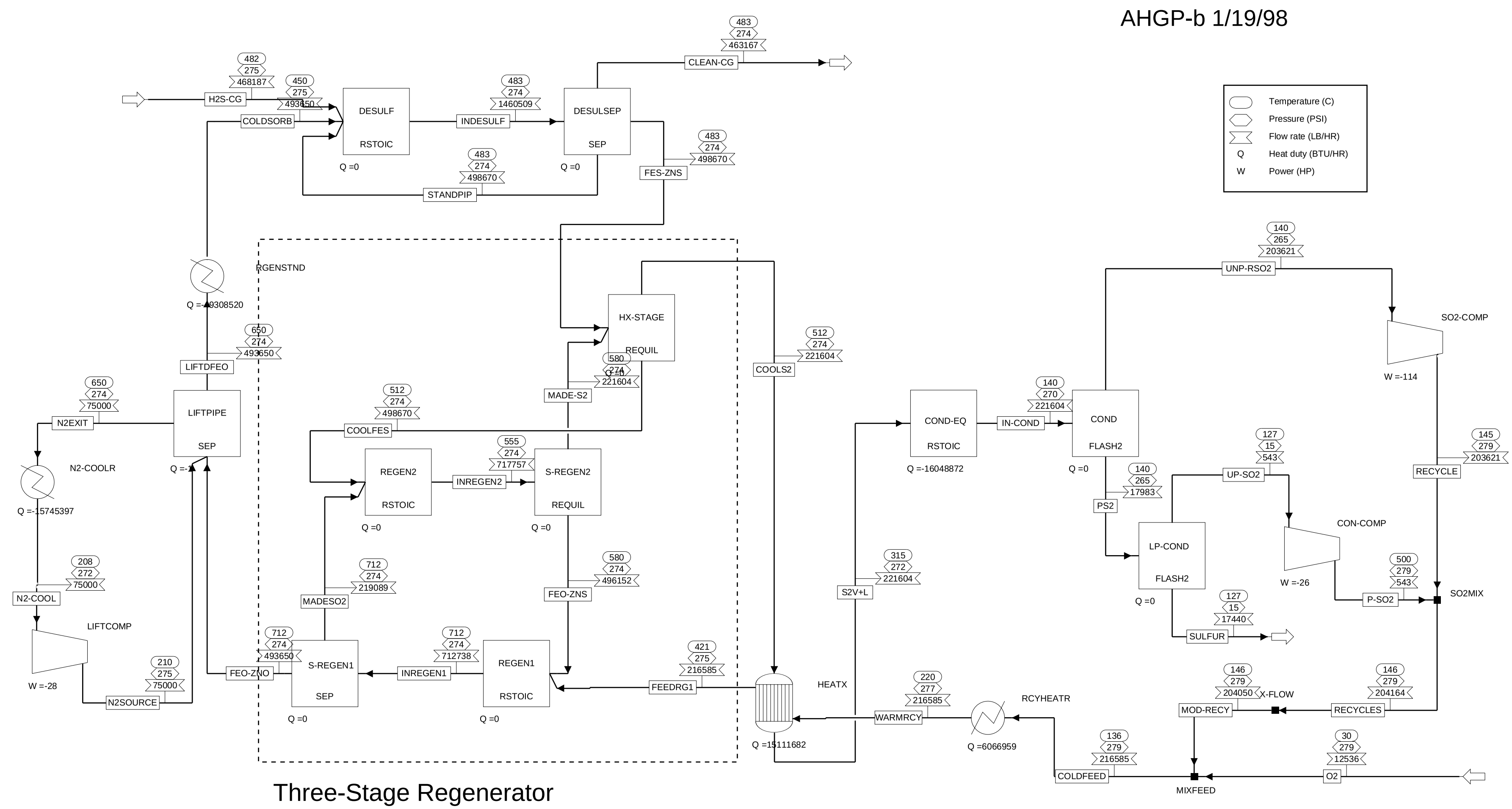
Display ALLS	TREAMS	P-SO2	PS2	RECYCLE	RECYCLES	S2V+L	STNDPIPE	SULFUR	UNP-RSO2	UP-SO2	WARMRCY		Display ALLS	TREAMS
Units:	From	CON-COMP	DEMISTR	SO2-COMP	SO2MIX	HEATX	DSULSTND	LP-COND	DEMISTR	LP-COND	RCYHEATR		Units:	From
Format: SOLI	DS To	SO2MIX	LP-COND	SO2MIX	X-FLOW	COND-EQ	DESULF		SO2-COMP	CON-COMP	HEATX		Format: SOLI	DS To
	Phas	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	MISSING	LIQUID	VAPOR	VAPOR	VAPOR			Phas
Temperature	[C]	500.6	139.6	145.1	146.1	315	482.4	127.1	139.6	127.1	240		Temperature	[C]
Pressure	[PSI]	279.2	265.7	279.2	279.2	272.7	274.7	14.7	265.7	14.7	277.2		Pressure	[PSI]
Mass VFrac		1	0	1	1	1	0	0	1	1	1		Mass VFrac	
Mass SFrac		0	0	0	0	0	1	0	0	0	0		Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	179.039	5910.173	68676.164	68855.203	74586.227	166009.094	5731.134	68676.133	179.039	72935.094		Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	148.551	33.732	27679.805	27840.887	44448.238	793.139	31.631	28725.764	1460.365	41204.621		Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-3.30E+05	120743.211	-1.35E+08	-135330000	-1.30E+08	-1.04E+09	472141.281	-135100000	-351400	-132710000		Enthalpy	[BTU/HR]
Density	[LB/CUFT]	1.205	175.208	2.481	2.473	1.678	209.307	181.187	2.391	0.123	1.77		Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S		178.945	190.34	68669.961	68848.906	68860.266		11.396	68669.93	178.945	68800		O2S	
H2S													H2S	
H2O													H2O	
S2						519.707							S2	
S6		0.036	2179.265	2.363	2.398	2171.238		2179.229	2.363	0.036	2.397		S6	
S8		0.058	3540.567	3.838	3.897	3035.012		3540.509	3.838	0.058	3.894		S8	
CO													CO	
CO2													CO2	
H2													H2	
O2											4128.805		O2	
N2													N2	
COS													COS	
ZNO													ZNO	
ZNS							2496.533						ZNS	
FE2O3													FE2O3	
FEO													FEO	
FES							13512.56						FES	
AL2O3							150000						AL2O3	

AHGP-b Advanced Hot Gas Process



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Display ALLS	TREAMS	CLEAN-CG	COLDFEED	COLDSORB	COOLFES	COOLS2	FEEDRG1	FEO-ZNO	FEO-ZNS	FES-ZNS	H2S-CG	IN-COND	Display ALLS	TREAMS
Units:	From	DESULSEP	MIXFEED	RGENSTND	HX-STAGE	HX-STAGE	HEATX	S-REGEN1	S-REGEN2	DESULSEP		COND-EQ	Units:	From
Format: SOLI	DS To		RCYHEATR	DESULF	REGEN2	HEATX	REGEN1	LIFTPIPE	REGEN1	HX-STAGE	DESULF	COND	Format: SOLI	DS To
	Phas	VAPOR	MIXED	MISSING	MISSING	VAPOR	VAPOR	MISSING	MISSING	MISSING	VAPOR	MIXED		Phas
Temperature	[C]	482.7	136.3	450	512.1	512.1	421.5	711.6	579.5	482.7	482.2	140	Temperature	[C]
Pressure	[PSI]	274.1	278.6	275	274.1	274.1	274.6	274.1	274.1	274.1	275	270.1	Pressure	[PSI]
Mass VFrac		1	1	0	0	1	1	0	0	0	1	0.919	Mass VFrac	
Mass SFrac		0	0	1	1	0	0	1	1	1	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	463167.313	216585.75	493649.531	498669.813	221603.984	216585.75	493649.531	496152.125	498669.813	468187.563	221603.984	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR	1.17E+06	91955.008	2270.583	2382.878	180074.938	173024.047	2270.583	2307.888	2382.878	1.16E+06	8.38E+04	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.16E+09	-4.01E+08	-3.18E+09	-3.10E+09	-3.69E+08	-3.80E+08	-3.12E+09	-3.13E+09	-3.11E+09	-1.08E+09	-4.00E+08	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.397	2.355	217.411	209.272	1.231	1.252	217.411	214.981	209.272	0.403	2.643	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			204031.219			204180.078	204031.219					204180.078	O2S	
H2S		23.471									18603.785		H2S	
H2O		83842.484									69811.359		H2O	
S2						1549.572							S2	
S6			7.099			6588.625	7.099					6619.617	S6	
S8			11.587			9285.715	11.587					10804.296	S8	
CO		215953.672									215953.672		CO	
CO2		116218.758									116218.758		CO2	
H2		11175.208									11646.276		H2	
O2			12535.85				12535.85						O2	
N2		35953.723									35953.723		N2	
COS													COS	
ZNO				6332.866				6332.866					ZNO	
ZNS					7583				7583	7583			ZNS	
FE2O3				37316.672				37316.672	24919.895				FE2O3	
FEO													FEO	
FES					41086.797				13649.23	41086.797			FES	
AL2O3				450000	450000			450000	450000	450000			AL2O3	

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Display ALLS	TREAMS	INDESULF	INREGEN1	INREGEN2	LIFTDFEO	MADE-S2	MADESO2	MOD-RECY	N2-COOL	N2EXIT	N2SOURCE	O2	Display ALLS	TREAMS
Units:	From	DESULF	REGEN1	REGEN2	LIFTPIPE	S-REGEN2	S-REGEN1	X-FLOW	N2-COOLR	LIFTPIPE	LIFTCOMP		Units:	From
Format: SOLI	DS To	DESULSEP	S-REGEN1	S-REGEN2	RGENSTND	HX-STAGE	REGEN2	MIXFEED	LIFTCOMP	N2-COOLR	LIFTPIPE	MIXFEED	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	482.7	711.6	554.9	649.8	579.5	711.6	146.1	208	649.8	210.1	30	Temperature	[C]
Pressure	[PSI]	274.1	274.1	274.1	274.1	274.1	274.1	278.6	272	274.1	275	278.6	Pressure	[PSI]
Mass VFrac		0.317	0.307	0.309	0	1	1	1	1	1	1	1	Mass VFrac	
Mass SFrac		0.683	0.693	0.691	1	0	0	0	0	0	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	1460510	712737.875	717757.063	493649.531	221603.984	219088.344	204049.906	75000	75000	75000	12535.85	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1171290	2.40E+05	202161.641	2270.583	198350.969	237827.297	82715.25	92161.805	175148.922	91557.102	8121.884	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-7.37E+09	-3.51E+09	-3.49E+09	-3.13E+09	-3.62E+08	-3.87E+08	-4.01E+08	6142950	2.19E+07	6213730	-5136.922	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	1.247	2.969	3.55	217.411	1.117	0.921	2.467	0.814	0.428	0.819	1.543	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			219000	204180.953		204180.078	219000	204031.219					O2S	
H2S		23.471											H2S	
H2O		83842.484											H2O	
S2				17423.979		4630.148							S2	
S6						6555.111		7.099					S6	
S8						6238.653		11.587					S8	
CO		215953.672											CO	
CO2		116218.758											CO2	
H2		11175.208											H2	
O2			88.344				88.344					12535.85	O2	
N2		35953.723							75000	75000	75000		N2	
COS													COS	
ZNO			6332.866		6332.866								ZNO	
ZNS		15166		7583									ZNS	
FE2O3			37316.672	24919.895	37316.672								FE2O3	
FEO													FEO	
FES		82173.594		13649.23									FES	
AL2O3		900000	450000	450000	450000								AL2O3	

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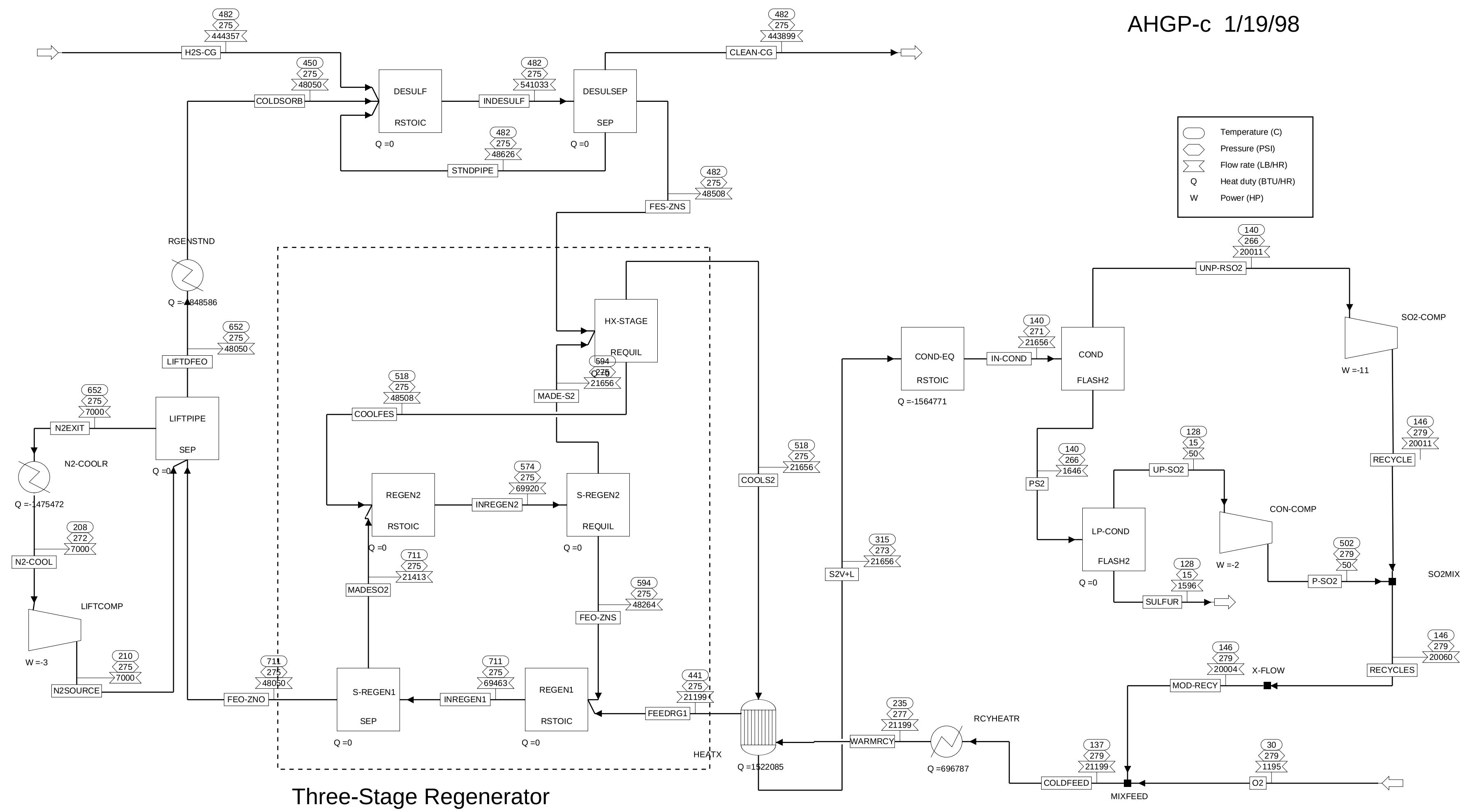
Display ALLS	TREAMS	P-SO2	PS2	RECYCLE	RECYCLES	S2V+L	STANDPIP	SULFUR	UNP-RSO2	UP-SO2	WARMRCY		Display ALLS	TREAMS
Units:	From	CON-COMP	COND	SO2-COMP	SO2MIX	HEATX	DESULSEP	LP-COND	COND	LP-COND	RCYHEATR		Units:	From
Format: SOLI	DS To	SO2MIX	LP-COND	SO2MIX	X-FLOW	COND-EQ	DESULF		SO2-COMP	CON-COMP	HEATX		Format: SOLI	DS To
	Phas	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	MISSING	LIQUID	VAPOR	VAPOR	VAPOR			Phas
Temperature	[C]	500.4	139.6	145.1	146.1	315	482.7	127.2	139.6	127.2	220		Temperature	[C]
Pressure	[PSI]	278.6	265.1	278.6	278.6	272.1	274.1	14.7	265.1	14.7	276.6		Pressure	[PSI]
Mass VFrac		1	0	1	1	1	0	0	1	1	1		Mass VFrac	
Mass SFrac		0	0	0	0	0	1	0	0	0	0		Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	543.2	17983.422	203620.531	204163.734	221603.984	498669.813	17440.221	203620.938	543.2	216585.75		Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	451.514	102.6	82271.875	82761.391	132160.859	2382.878	96.225	85387.891	4431.133	117188.945		Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.00E+06	370562.063	-4.00E+08	-401270000	-3.84E+08	-3.11E+09	1436680	-400560000	-1066100	-394990000		Enthalpy	[BTU/HR]
Density	[LB/CUFT]	1.203	175.276	2.475	2.467	1.677	209.272	181.244	2.385	0.123	1.848		Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S		542.913	577.553	203602.125	204145.031	204180.078		34.64	203602.531	542.913	204031.219		O2S	
H2S													H2S	
H2O													H2O	
S2						1549.572							S2	
S6		0.109	6612.687	6.994	7.103	6588.625		6612.578	6.994	0.109	7.099		S6	
S8		0.178	10793.182	11.415	11.593	9285.715		10793.004	11.415	0.178	11.587		S8	
CO													CO	
CO2													CO2	
H2													H2	
O2											12535.85		O2	
N2													N2	
COS													COS	
ZNO													ZNO	
ZNS							7583						ZNS	
FE2O3													FE2O3	
FEO													FEO	
FES							41086.797						FES	
AL2O3							450000						AL2O3	

AHGP-c Advanced Hot Gas Process

AHGP-c 1/19/98



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Display ALLS	TREAMS	CLEAN-CG	COLDFEED	COLDSORB	COOLFES	COOLS2	FEEDRG1	FEO-ZNO	FEO-ZNS	FES-ZNS	H2S-CG	IN-COND	Display ALLS	TREAMS
Units:	From	DESULSEP	MIXFEED	RGENSTND	HX-STAGE	HX-STAGE	HEATX	S-REGEN1	S-REGEN2	DESULSEP		COND-EQ	Units:	From
Format: SOLI	DS To		RCYHEATR	DESULF	REGEN2	HEATX	REGEN1	LIFTPIPE	REGEN1	HX-STAGE	DESULF	DEMISTR	Format: SOLI	DS To
	Phas	VAPOR	MIXED	MISSING	MISSING	VAPOR	VAPOR	MISSING	MISSING	MISSING	VAPOR	MIXED		Phas
Temperature	[C]	482.1	136.9	450	517.8	517.8	441.4	711	594.5	482.1	482.2	140	Temperature	[C]
Pressure	[PSI]	274.9	279.4	275	274.9	274.9	275.4	275.4	275.4	274.9	275	270.9	Pressure	[PSI]
Mass VFrac		1	1	0	0	1	1	0	0	0	1	0.924	Mass VFrac	
Mass SFrac		0	0	1	1	0	0	1	1	1	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	443898.906	21199.451	48050.277	48507.699	21656.5	21199.451	48050.277	48263.68	48507.699	444356.313	21656.5	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1.12E+06	8971.891	221.013	231.696	17760.699	17373.199	221.013	224.428	231.696	1.12E+06	8.21E+03	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.07E+09	-3.93E+07	-3.11E+08	-3.03E+08	-3.62E+07	-3.71E+07	-3.05E+08	-3.05E+08	-3.04E+08	-1.06E+09	-3.93E+07	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.397	2.363	217.409	209.359	1.219	1.22	217.409	215.052	209.359	0.397	2.638	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			20002.434			20061.83	20002.434					20061.83	O2S	
H2S		65.815									1790.604		H2S	
H2O		70033.773									68721.594		H2O	
S2						164.784							S2	
S6			0.719			612.014	0.719					615.31	S6	
S8			1.144			817.872	1.144					979.36	S8	
CO		212582.609									212582.609		CO	
CO2		114404.563									114404.563		CO2	
H2		11419.664									11464.477		H2	
O2			1195.155				1195.155						O2	
N2		35392.48									35392.48		N2	
COS													COS	
ZNO				500.379				500.379					ZNO	
ZNS					599.156				599.156	599.156			ZNS	
FE2O3				3549.896				3549.896	2415.33				FE2O3	
FEO													FEO	
FES					3908.545				1249.193	3908.545			FES	
AL2O3				44000	44000			44000	44000	44000			AL2O3	

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Display ALLS	TREAMS	INDESULF	INREGEN1	INREGEN2	LIFTDFEO	MADE-S2	MADESO2	MOD-RECY	N2-COOL	N2EXIT	N2SOURCE	O2	Display ALLS	TREAMS
Units:	From	DESULF	REGEN1	REGEN2	LIFTPIPE	S-REGEN2	S-REGEN1	X-FLOW	N2-COOLR	LIFTPIPE	LIFTCOMP		Units:	From
Format: SOLI	DS To	DESULSEP	S-REGEN1	S-REGEN2	RGENSTND	HX-STAGE	REGEN2	MIXFEED	LIFTCOMP	N2-COOLR	LIFTPIPE	MIXFEED	Format: SOLI	DS To
	Phas	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR		Phas
Temperature	[C]	482.1	711	573.5	651.5	594.5	711	146.5	208	651.5	210.1	30	Temperature	[C]
Pressure	[PSI]	274.9	275.4	275.4	275	275.4	275.4	279.4	272	275	275	279.4	Pressure	[PSI]
Mass VFrac		0.82	0.308	0.31	0	1	1	1	1	1	1	1	Mass VFrac	
Mass SFrac		0.18	0.692	0.69	1	0	0	0	0	0	0	0	Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	541032.563	69463.133	69920.453	48050.277	21656.5	21412.854	20004.297	7000	7000	7000	1195.155	Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	1119290	2.34E+04	20135.34	221.013	19779.051	23221.412	8092.941	8601.769	16324.121	8545.33	772.086	Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-1.68E+09	-3.42E+08	-3.41E+08	-3.06E+08	-3.54E+07	-3.76E+07	-3.93E+07	573341.625	2.05E+06	579948.125	-497.809	Enthalpy	[BTU/HR]
Density	[LB/CUFT]	0.483	2.963	3.473	217.409	1.095	0.922	2.472	0.814	0.429	0.819	1.548	Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S			21310.348	20062.115		20061.83	21310.348	20002.434					O2S	
H2S		65.815											H2S	
H2O		70033.773											H2O	
S2				1594.662		542.278							S2	
S6						569.93		0.719					S6	
S8						482.462		1.144					S8	
CO		212582.609											CO	
CO2		114404.563											CO2	
H2		11419.664											H2	
O2			102.507				102.507					1195.155	O2	
N2		35392.48							7000	7000	7000		N2	
COS													COS	
ZNO			500.379		500.379								ZNO	
ZNS		1316.587		599.156									ZNS	
FE2O3			3549.896	2415.33	3549.896								FE2O3	
FEO													FEO	
FES		7817.09		1249.193									FES	
AL2O3		88000	44000	44000	44000								AL2O3	

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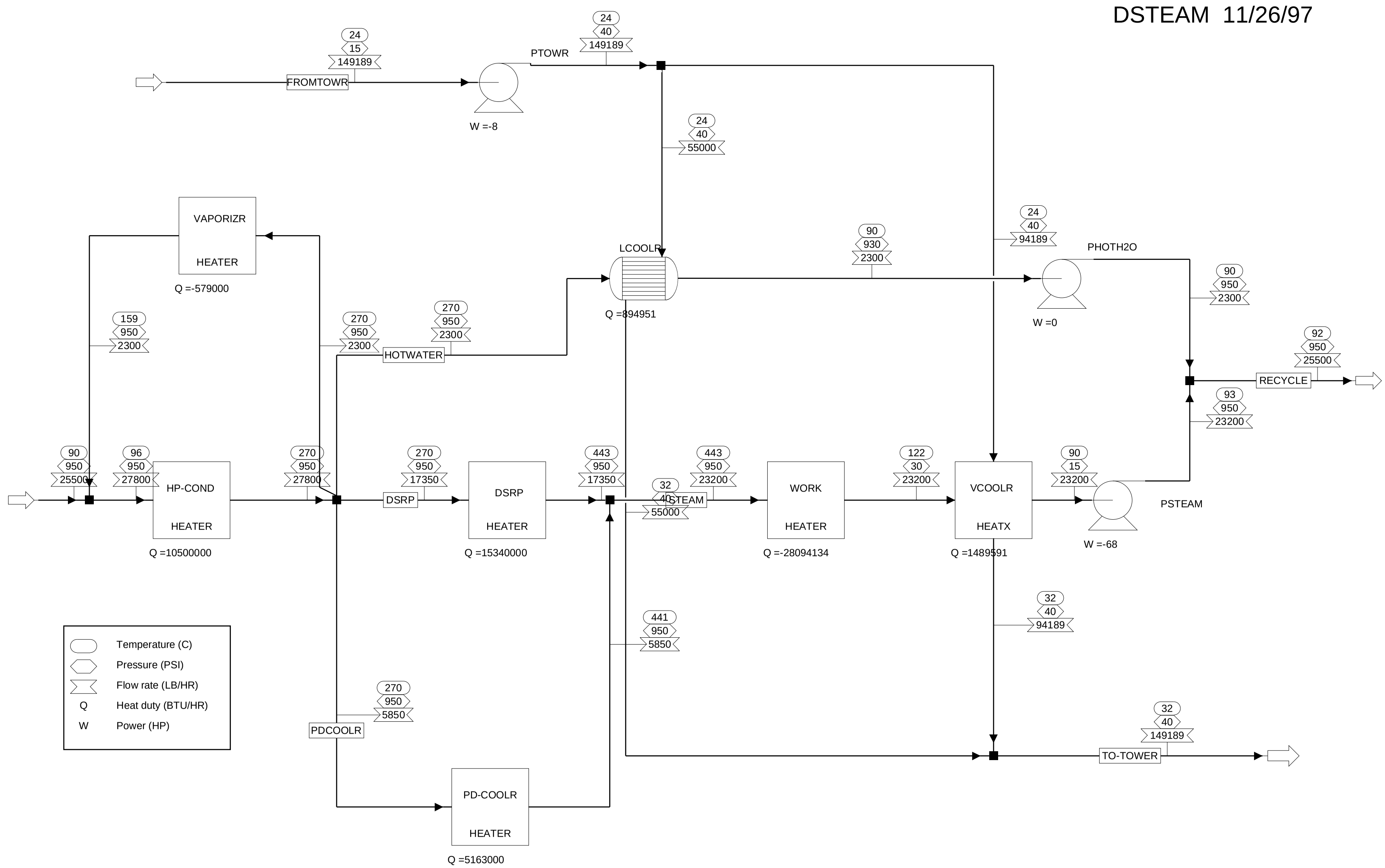
Display ALLS	TREAMS	P-SO2	PS2	RECYCLE	RECYCLES	S2V+L	STNDPIPE	SULFUR	UNP-RSO2	UP-SO2	WARMRCY		Display ALLS	TREAMS
Units:	From	CON-COMP	DEMISTR	SO2-COMP	SO2MIX	HEATX	DESULSEP	LP-COND	DEMISTR	LP-COND	RCYHEATR		Units:	From
Format: SOLI	DS To	SO2MIX	LP-COND	SO2MIX	X-FLOW	COND-EQ	DESULF		SO2-COMP	CON-COMP	HEATX		Format: SOLI	DS To
	Phas	VAPOR	LIQUID	VAPOR	VAPOR	VAPOR	MISSING	LIQUID	VAPOR	VAPOR	VAPOR			Phas
Temperature	[C]	501.6	140	145.5	146.5	315	482.1	127.7	140	127.7	235		Temperature	[C]
Pressure	[PSI]	279.4	265.9	279.4	279.4	272.9	274.9	14.7	265.9	14.7	277.4		Pressure	[PSI]
Mass VFrac		1	0	1	1	1	0	0	1	1	1		Mass VFrac	
Mass SFrac		0	0	0	0	0	1	0	0	0	0		Mass SFrac	
*** ALL PHAS	ES ***												*** ALL PHAS	ES ***
Mass Flow	[LB/HR]	49.58	1645.675	20010.68	20060.26	21656.5	48625.977	1596.094	20010.973	49.58	21199.451		Mass Flow	[LB/HR]
Volume Flow	[CUFT/HR]	41.161	9.405	8070.953	8115.582	12934.794	232.279	8.823	8375.842	404.953	11828.04		Volume Flow	[CUFT/HR]
Enthalpy	[BTU/HR]	-9.14E+04	34663.465	-3.93E+07	-39425000	-3.78E+07	-3.04E+08	131964.156	-39363000	-97300.727	-38619000		Enthalpy	[BTU/HR]
Density	[LB/CUFT]	1.205	174.979	2.479	2.472	1.674	209.343	180.904	2.389	0.122	1.792		Density	[LB/CUFT]
Mass Flow	[LB/HR]												Mass Flow	[LB/HR]
O2S		49.553	52.704	20008.838	20058.391	20061.83		3.15	20009.131	49.553	20002.434		O2S	
H2S													H2S	
H2O													H2O	
S2						164.784							S2	
S6		0.01	614.651	0.711	0.721	612.014		614.641	0.711	0.01	0.719		S6	
S8		0.016	978.32	1.131	1.148	817.872		978.303	1.131	0.016	1.144		S8	
CO													CO	
CO2													CO2	
H2													H2	
O2											1195.155		O2	
N2													N2	
COS													COS	
ZNO													ZNO	
ZNS							717.431						ZNS	
FE2O3													FE2O3	
FEO													FEO	
FES							3908.545						FES	
AL2O3							44000						AL2O3	

Appendix F
Steam Generation Process Flowsheets

The following flowsheets represent possible design schemes for producing high pressure steam. Desulfurization units that require heat removal are utilized for producing the steam. The steam generated will result in an economic credit for the process. The steam generation simulations will help determine the equipment necessary for cooling the desulfurization process.

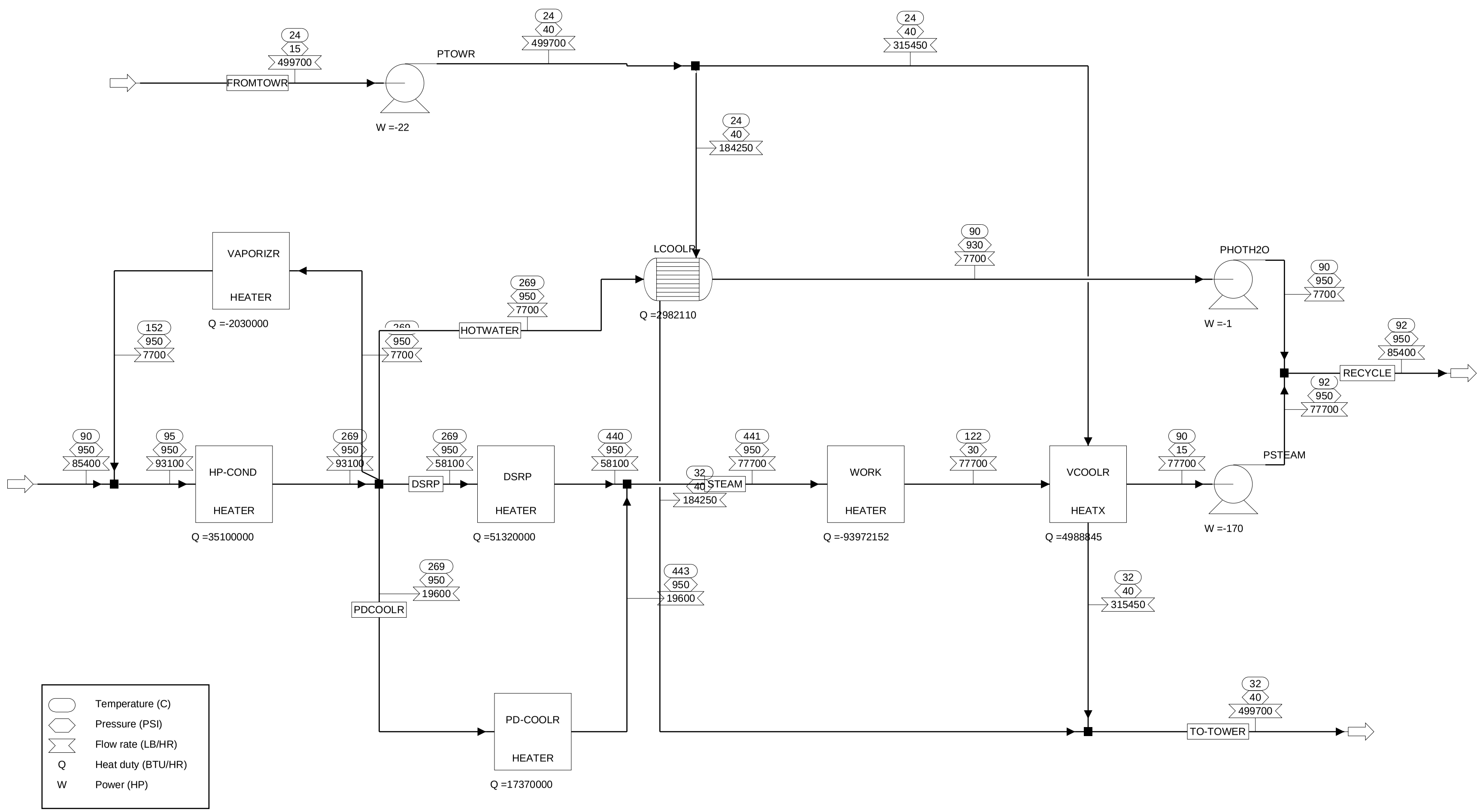
DSRP Complete Steam Generation Scheme

DSTEAM 11/26/97



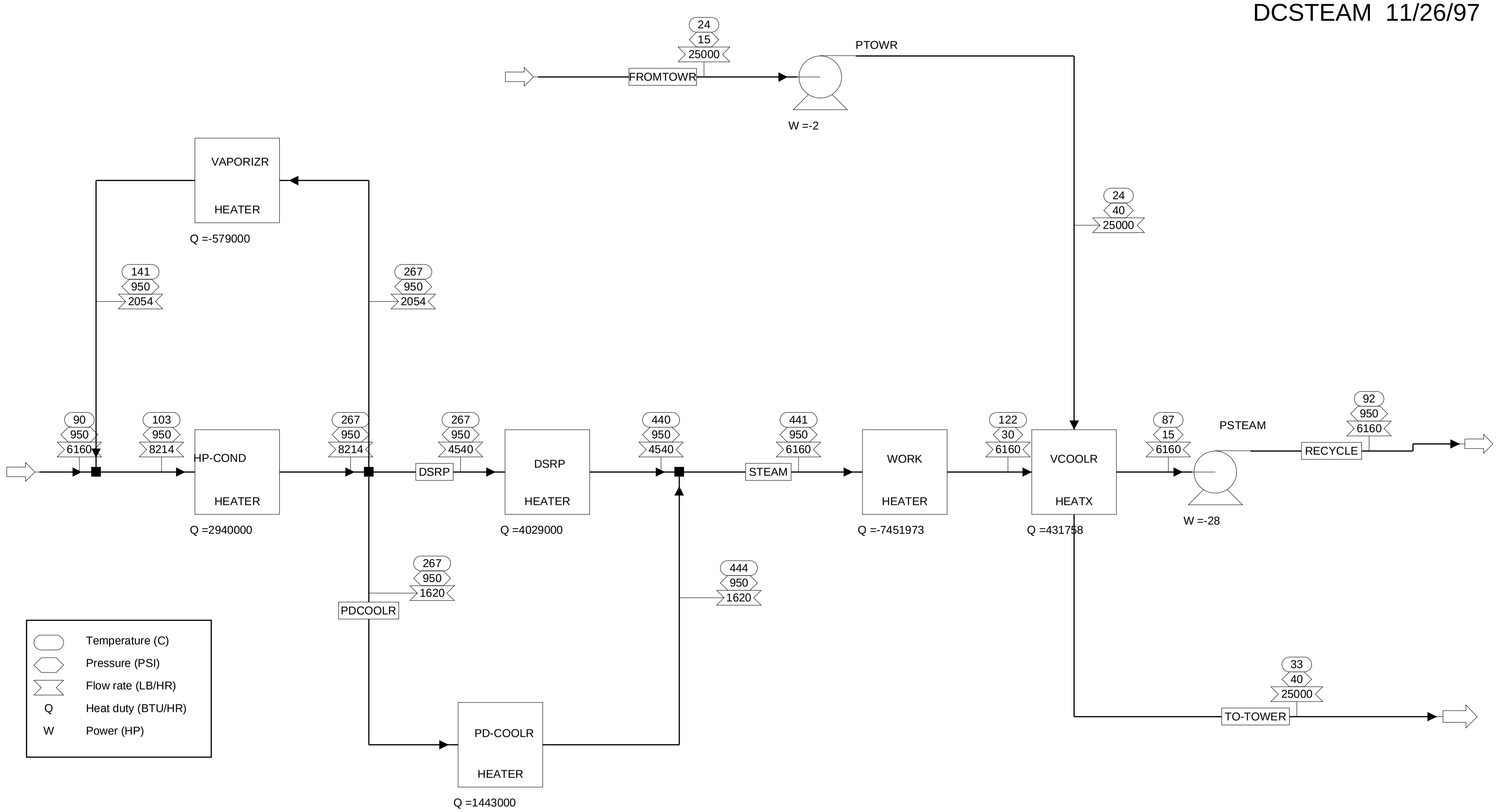
DSRP-b Complete Steam Generation Scheme

DBSTEAM 11/26/97



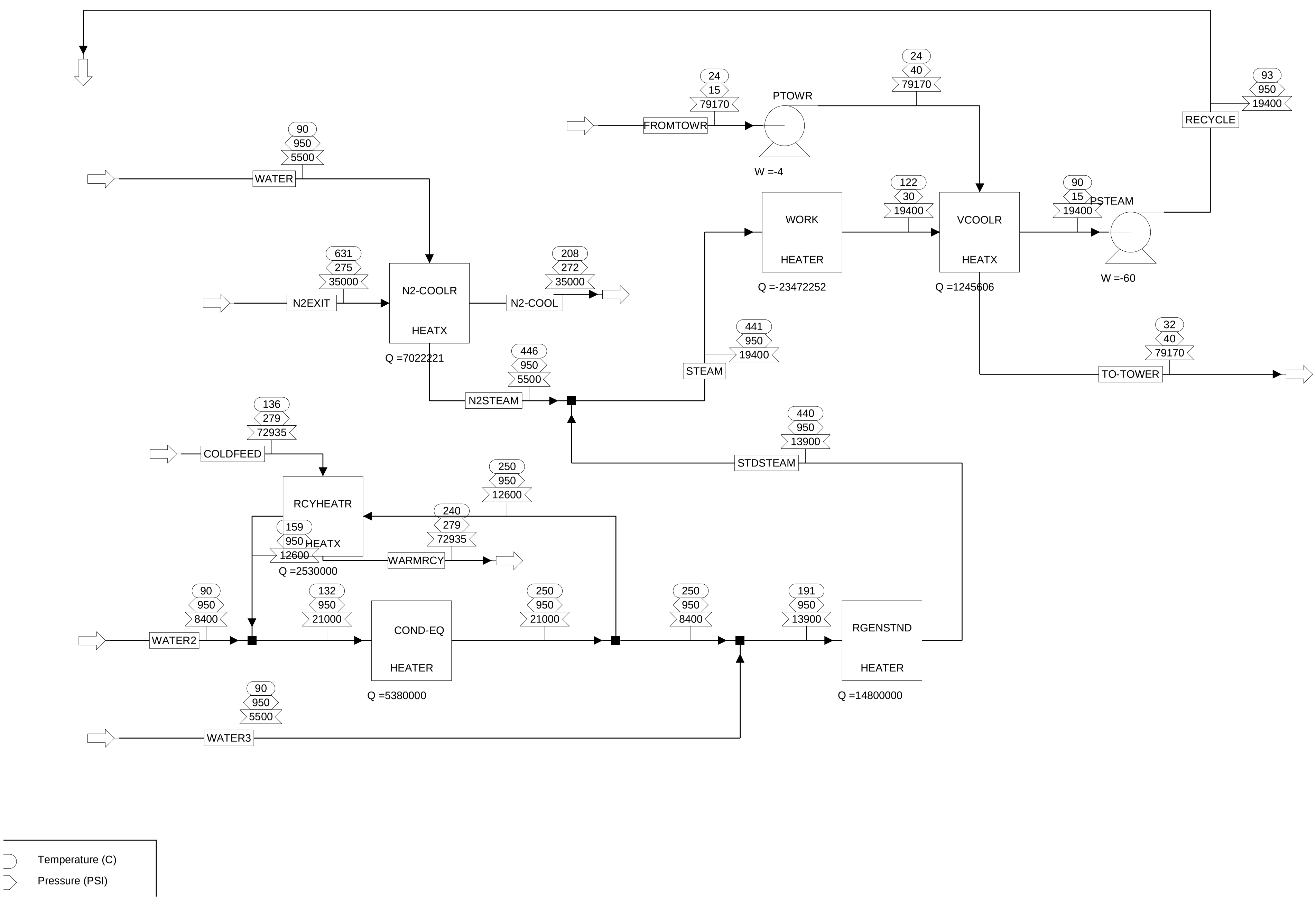
DSRP-c Complete Steam Generation Scheme

DCSTEAM 11/26/97



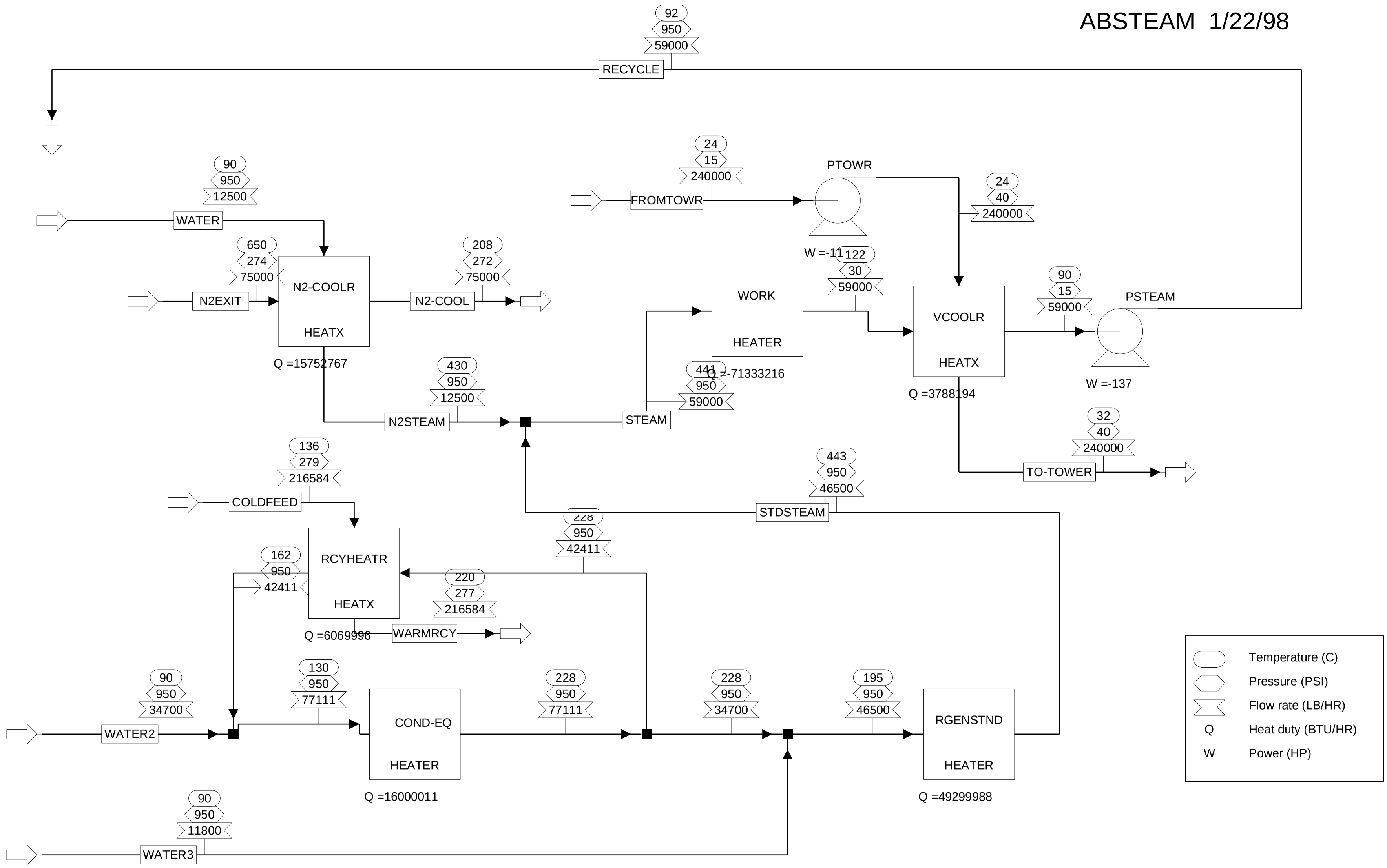
AHGP Complete Steam Generation Scheme

ASTEAM 1/22/98



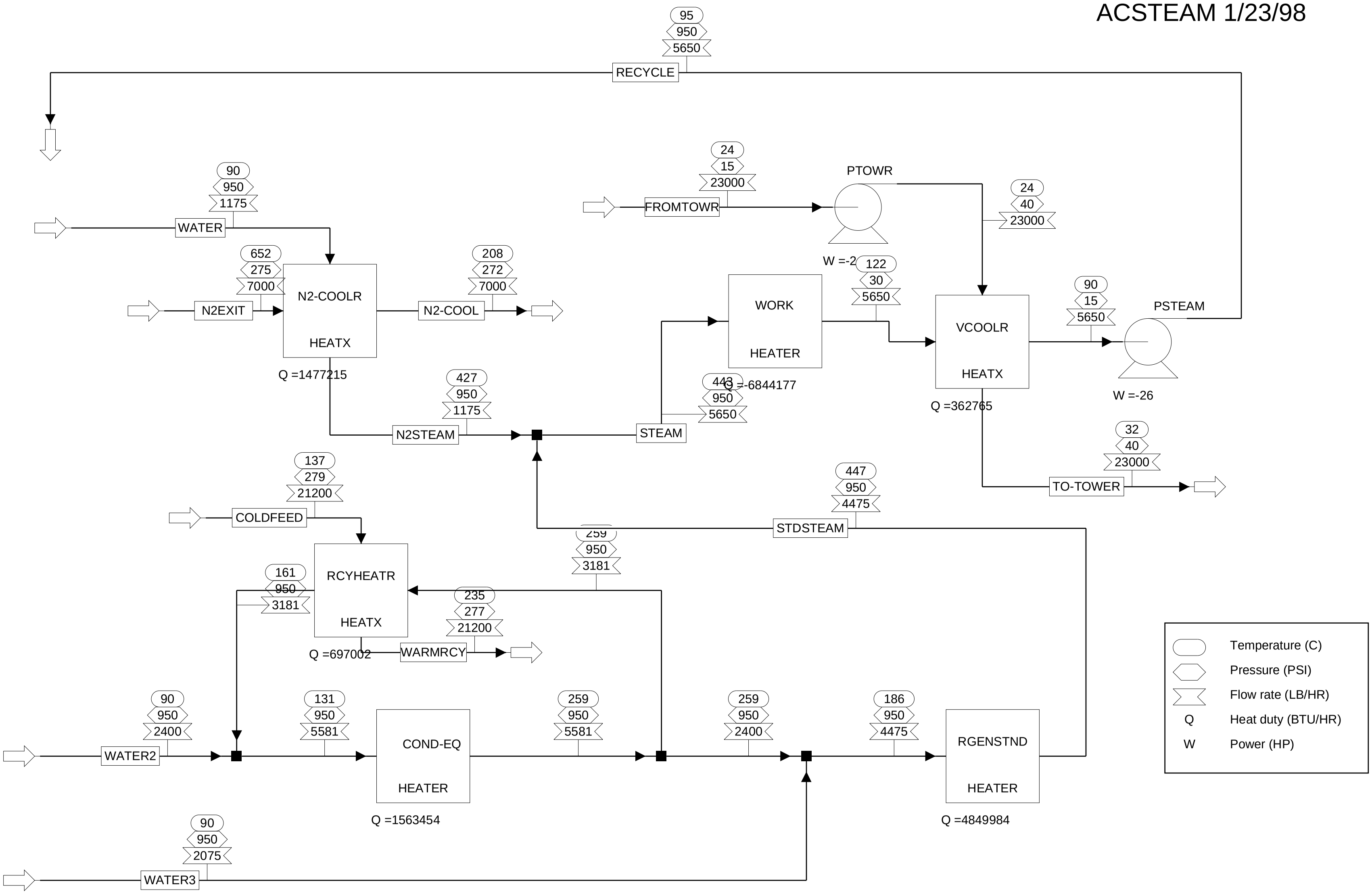
AHGP-b Complete Steam Generation Scheme

ABSTEAM 1/22/98



AHGP-c Complete Steam Generation Scheme

ACSTEAM 1/23/98



Appendix G
Calculation of Reactor Size

The reactor's diameter is determined from the average volumetric flow rate and the linear velocity.

v = 20 ft/sec = 72,000 ft/hr

V = [(gas volume entering) + (gas volume leaving)]/2 + (sorbent mass flow) $(60\ lb/ft^3)^{-1}$

Area = V/v = (Volumetric flow rate ft^3/hr) / 72,000 ft/hr {ft^2}

Calculating the area allows for the calculation of the reactor inside diameter.

Area = π $(I.D.)^2$ / (4 x 144 in^2/ft^2)

I.D. = $[\ (Area)\ (4 \times 144) / \pi\ ]^{0.5}$ {in}

The reactor cost will be based on the material of construction costs. The reactor wall thickness and height are necessary for such a calculation. The reactor system cost will be calculated to include installation costs.

The reactor will be cylindrical. The wall and heads will be assumed to have the same thickness. The following equation was used for determining wall thickness (Peters & Timmerhaus, 1991).

Thickness = P (I.D.) / [2 (Max. allowable working stress psia) (Efficiency of joints) - 0.6 P] + C_c

P = pressure {psia} C_c = corrosion losses {in}

Thickness = 275 (I.D.)/ [2 (12,000) (0.85) - 0.6 (275)] + 0.125 {in}

Taking steel density to be 489 lb/ft^3, the reactor weight is calculated with the equations below.

Weight of shell = π (I.D./12) (height) (Thickness/12)(489) {lbs}

Weight of heads = 2 π $[12\ I.D./\ 2]^2$ (Thickness) $(489/12^3)$ (2) {lbs}

Total weight = 1.15 (weight of shell + weight of heads) {lbs}

The total weight is increased 15% to account of nozzles, manholes, ect.

The cost of carbon steel can be calculated by the equation below.

$$(\text{Cost per lb}) = 80\ (\text{total weight})^{-0.34}$$

The equation above in applicable for 800 lb to 100,000 lb vessels (Peters and Timmerhaus 1991). Estimates for weights over 100,000 lbs could not be found. Therefore, in such cases the unit cost for carbon steel was taken as an average of the above equation calculated for 100,000 lb and the above equation calculated for the total weight. The unit cost is expected to continue to decrease at larger quantities but the decrease should become less pronounced.

Unit cost of carbon steel (weight > 100,000 lbs)

$$(\text{Cost per lb}) = 80\ [(\text{total weight})^{-0.34} + (100{,}000)^{-0.34}]\ /\ 2$$

The cost of installation will be twice of the cost of the reactor if it were constructed of carbon steel.

$$(\text{Cost of installation}) = 2\ (\text{Cost per lb})\ (\text{total weight})$$

The total cost of the reactor system includes installation and material costs. Material cost is multiplied by 3.5 to account for using stainless steel 310 instead of carbon steel.

$$(\text{Total cost for reactor}) = (\text{Cost of installation}) + 3.5\ (\text{Cost per lb})\ (\text{total weight})$$

Appendix H
Sizing Reactors for the DSRP

Copies of the reactor system sizing calculations follow. They include estimates of the reactor system costs. The equations describe in *Appendix G - Calculation of Reactor Size* where used in the spreadsheet.

Desulf and Regen transport reactor price calculation

DSRP

Regenerator Reactor

v (ft/sec) =	20		72000	ft/hr	air volume	67,895	cfh	HP-O2-N2
V (cfh) =	85,541				ROG volume	94,813	cfh	ROG
Area =	1.188	ft^2			regen sorbent flow	251,240	lb/hr	ZNS2RGEN
I.D.	15.068	in			regen sorbent vol.	4,187	cfh	
					sorbent vol%	4.90%		
thickness=	0.330							
shell wt.=	5,302	lbs			Corrosion depth	0.125	in	
heads wt.=	67	lbs			reactor height	100	ft	

total wt. 6,174 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

size vs. regen size 1

total wt. 6,174 lbs

Desulfurization Reactor

v (ft/sec) =	20		72000	ft/hr	coal gas in volume	1,200,000	cfh	RAW-CG
V (cfh) =	1,296,166				cg out volume	1,370,000	cfh	CG-CALC
Area =	18.002	ft^2			regen sorbent flow	669,972	lb/hr	ZNS
I.D.	58.653	in			regen sorbent vol.	11,166	cfh	
					sorbent vol%	0.86%		
thickness=	0.922							
shell wt.=	57,707	lbs			Corrosion depth	0.125	in	
heads wt.=	2,821	lbs			reactor height	100	ft	

total wt. 69,607 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

size vs. desulf size 1

total wt. 69,607 lbs

total wt. 151,561 lbs weight for desulfurization and regeneration transport reactors

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.491	$/lb	1990 $	1.490988	1.386
	1.593	$/lb	1996 $		<100,000 lb calc

Cost of installation $482,917

Total reactor cost **$1,328,020** includes cost of installation

DSRP Reactor Cost

DSRP

DSRP Reactor

v (ft/sec) gas =	3	10800	ft/hr	slipstream	37,342	cfh	SLIPSTREAM
V (cfh) =	114,923			ROG volume	75,166	cfh	ROG-COOL
space time -gas	33.33	seconds		reactor effluent	107,359	cfh	RXNPRD
v (ft/sec) cat =	2.3	8280	ft/hr	DSRP reactor Q	-15,340,000	BTU/hr	
				catalyst flow	299,381	lb/hr	
				catalyst vol.	4,990	cfh	
Area =	10.782	ft^2					
I.D.	45.391	in		catalyst vol%	5.59%		
thickness=	0.742	in					
shell wt.=	35,930	lbs		Corrosion depth	0.125	in	
heads wt.=	1,359	lbs		reactor height	100	ft	

total wt. 42,882 lbs (includes additional 15% for nozzles, manholes, etc.)

DSRP Standpipe

Cyclone (20% of reactor size)	8,576		Area =	10.78	ft^2
			I.D.	45.39	in
standpipe height	40	ft	thickness=	0.74	in
			shell wt.=	14,372	lbs
residense time	10.81	minutes	heads wt.=	1,359	lbs

total wt. 26,667 lbs (includes additional 15% on standpipe weight + Cyclone weight)

Heat Exchanger

Heat Exchanger Area (ft^2) 1063 heat exchanger pipe thickness 0.25 in

volume of steel 22

total weight 10,829 lbs

total wt. 80,379 lbs weight for DSRP reactor system

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.719	$/lb	1990 $	1.657735955	1.719
	1.837	$/lb	1996 $		<100,000 lb calc

Cost of installation $295,320

Total reactor cost $812,129 includes cost of installation

Desulf and Regen transport reactor price calculation

DSRP

Regenerator Reactor					air volume	215,340	cfh	HP-O2-N2
v (ft/sec) =	20		72000	ft/hr	ROG volume	299,541	cfh	ROG
V (cfh) =	277,458				regen sorbent flow	1,201,050	lb/hr	ZNS2RGEN
Area =	3.854	ft^2			regen sorbent vol.	20,018	cfh	
I.D.	27.137	in						
					sorbent vol%	7.21%		
thickness=	0.494							
shell wt.=	14,298	lbs			Corrosion depth	0.125	in	
heads wt.=	323	lbs			reactor height	100	ft	

total wt. 16,814 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

size vs. regen size 1

total wt. 16,814 lbs

Desulfurization Reactor					coal gas in volume	1,409,220	cfh	RAW-CG
v (ft/sec) =	20		72000	ft/hr	cg out volume	1,877,650	cfh	CG-CALC
V (cfh) =	1,687,918				regen sorbent flow	2,669,000	lb/hr	ZNS
Area =	23.443	ft^2			regen sorbent vol.	44,483	cfh	
I.D.	66.933	in						
					sorbent vol%	2.64%		
thickness=	1.035							
shell wt.=	73,889	lbs			Corrosion depth	0.125	in	
heads wt.=	4,121	lbs			reactor height	100	ft	

total wt. 89,711 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

size vs. desulf size 1

total wt. 89,711 lbs

total wt. 213,051 lbs weight for desulfurization and regeneration transport reactors

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.415	$/lb	1990 $	1.4152335	1.234
	1.512	$/lb	1996 $		<100,000 lb calc

Cost of installation $644,349

Total reactor cost $1,771,959 includes cost of installation

DSRP-b Reactor Cost

DSRP

DSRP Reactor

v (ft/sec) gas =	3	10800	ft/hr	slipstream	153,708	cfh	SLIPSTREAM
V (cfh) =	393,089			ROG volume	237,184	cfh	ROG-COOL
space time -gas	33.33 seconds			reactor effluent	361,900	cfh	RXNPRD
v (ft/sec) cat =	2.3	8280	ft/hr	DSRP reactor Q	-51,320,000	BTU/hr	
				catalyst flow	1,001,581	lb/hr	
				catalyst vol.	16,693	cfh	
Area =	36.868 ft^2						
I.D.	83.936 in			catalyst vol%	5.47%		
thickness=	1.266 in						
shell wt.=	113,355 lbs			Corrosion depth	0.125	in	
heads wt.=	7,929 lbs			reactor height	100	ft	

total wt. 139,477 lbs (includes additional 15% for nozzles, manholes, etc.)

DSRP Standpipe

Cyclone (20% of reactor size)		27,895	Area =	36.87 ft^2
			I.D.	83.94 in
standpipe height	40 ft		thickness=	1.27 in
			shell wt.=	45,342 lbs
residense time	11.05 minutes		heads wt.=	7,929 lbs

total wt. 89,157 lbs (includes additional 15% on standpipe weight + Cyclone weight)

Heat Exchanger

Heat Exchanger Area (ft^2) 3556 heat exchanger pipe thickness 0.25 in

volume of steel 74

total weight 36,229 lbs

total wt. 264,863 lbs weight for DSRP reactor system

COST

			>100,000 lb calc	
C.S. unit price for quantiy needed	1.371 $/lb	1990 $	1.371208108	1.146
	1.465 $/lb	1996 $		<100,000 lb calc

Cost of installation $776,129

Total reactor cost $2,134,355 includes cost of installation

Desulf and Regen transport reactor price calculation

DSRP

Regenerator Reactor

v (ft/sec) =	20		72000 ft/hr	air volume	19,366	cfh	HP-O2-N2
V (cfh) =	23,882			ROG volume	26,009	cfh	ROG
Area =	0.332	ft^2		regen sorbent flow	71,663	lb/hr	ZNS2RGEN
I.D.	7.962	in		regen sorbent vol.	1,194	cfh	
				sorbent vol%	5.00%		
thickness=	0.233						
shell wt.=	1,981	lbs		Corrosion depth	0.125	in	
heads wt.=	13	lbs		reactor height	100	ft	

total wt. 2,293 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

size vs. regen size 1

total wt. 2,293 lbs

Desulfurization Reactor

v (ft/sec) =	20		72000 ft/hr	coal gas in volume	1,139,050	cfh	RAW-CG
V (cfh) =	1,164,094			cg out volume	1,186,750	cfh	CG-CALC
Area =	16.168	ft^2		regen sorbent flow	71,663	lb/hr	ZNS
I.D.	55.585	in		regen sorbent vol.	1,194	cfh	
				sorbent vol%	0.10%		
thickness=	0.880						
shell wt.=	52,215	lbs		Corrosion depth	0.125	in	
heads wt.=	2,419	lbs		reactor height	100	ft	

total wt. 62,829 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

size vs. desulf size 1

total wt. 62,829 lbs

total wt. 130,244 lbs weight for desulfurization and regeneration transport reactors

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.528	$/lb	1990 $	1.5276336	1.459
	1.632	$/lb	1996 $		<100,000 lb calc

Cost of installation $425,193

Total reactor cost $1,169,282 includes cost of installation

DSRP-c Reactor Cost

DSRP

DSRP Reactor

v (ft/sec) gas =	3	10800	ft/hr	slipstream	9,443	cfh	SLIPSTREAM
V (cfh) =	31,212			ROG volume	20,364	cfh	ROG-COOL
space time -gas	33.33 seconds			reactor effluent	29,995	cfh	RXNPRD
v (ft/sec) cat =	2.3	8280	ft/hr	DSRP reactor Q	-4,029,000	BTU/hr	
				catalyst flow	78,632	lb/hr	
				catalyst vol.	1,311	cfh	
Area =	2.93 ft^2						
I.D.	23.650 in			catalyst vol%	5.41%		
thickness=	0.446 in						
shell wt.=	11,265 lbs			Corrosion depth	0.125	in	
heads wt.=	222 lbs			reactor height	100	ft	

total wt. 13,210 lbs (includes additional 15% for nozzles, manholes, etc.)

DSRP Standpipe

Cyclone (20% of reactor size)	2,642	Area =	2.93	ft^2
		I.D.	23.65	in
standpipe height	40 ft	thickness=	0.45	in
		shell wt.=	4,506	lbs
residense time	11.17 minutes	heads wt.=	222	lbs

total wt. 8,079 lbs (includes additional 15% on standpipe weight + Cyclone weight)

Heat Exchanger

Heat Exchanger Area (ft^2) 279 heat exchanger pipe thickness 0.25 in

volume of steel 6

total weight 2,844 lbs

total wt. **24,133 lbs** weight for DSRP reactor system

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	2.588	$/lb	1990 $	2.092218539	2.588
	2.766	$/lb	1996 $		<100,000 lb calc

Cost of installation $133,482

Total reactor cost **$367,075** includes cost of installation

Desulf and Regen transport reactor price calculation

DSRP

Regenerator Reactor

v (ft/sec) =	20		72000	ft/hr	air volume	28,592	cfh	HP-O2-N2
V (cfh) =	36,020				ROG volume	39,921	cfh	ROG
Area =	0.500	ft^2			regen sorbent flow	105,797	lb/hr	ZNS2RGEN
I.D.	9.778	in			regen sorbent vol.	1,763	cfh	
					sorbent vol%	4.90%		
thickness=	0.258							
shell wt.=	2,690	lbs			Corrosion depth	0.125	in	
heads wt.=	22	lbs			reactor height	100	ft	

total wt. 3,119 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

size vs. regen size 1

total wt. 3,119 lbs

Desulfurization Reactor

v (ft/sec) =	20		72000	ft/hr	coal gas in volume	506,745	cfh	RAW-CG
V (cfh) =	545,644				cg out volume	575,144	cfh	CG-CALC
Area =	7.58	ft^2			regen sorbent flow	281,971	lb/hr	ZNS
I.D.	38.06	in			regen sorbent vol.	4,700	cfh	
					sorbent vol%	0.86%		
thickness=	0.642							
shell wt.=	26,075	lbs			Corrosion depth	0.125	in	
heads wt.=	827	lbs			reactor height	100	ft	

total wt. 30,938 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

size vs. desulf size 1

total wt. 30,938 lbs

total wt. 68,113 lbs weight for desulfurization and regeneration transport reactors

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.819	$/lb	1990 $	1.7075181	1.819
	1.943	$/lb	1996 $		<100,000 lb calc

Cost of installation $264,748

Total reactor cost $728,057 includes cost of installation

DSRP-100 Reactor Cost

DSRP

DSRP Reactor

v (ft/sec) gas =	3		10800 ft/hr	slipstream	15,723	cfh	SLIPSTREAM
V (cfh) =	48,391			ROG volume	31,647	cfh	ROG-COOL
space time -gas	33.33	seconds		reactor effluent	45,210	cfh	RXNPRD
v (ft/sec) cat =	2.3		8280 ft/hr	DSRP reactor Q	-6,459,000	BTU/hr	
				catalyst flow	126,056	lb/hr	
				catalyst vol.	2,101	cfh	
Area =	4.540	ft^2					
I.D.	29.454	in		catalyst vol%	5.59%		
thickness=	0.525	in					
shell wt.=	16,508	lbs		Corrosion depth	0.125	in	
heads wt.=	405	lbs		reactor height	100	ft	

total wt. 19,451 lbs (includes additional 15% for nozzles, manholes, etc.)

DSRP Standpipe

Cyclone (20% of reactor size)		3,890	Area =	4.54	ft^2
			I.D.	29.45	in
standpipe height	40 ft		thickness=	0.53	in
			shell wt.=	6,603	lbs
residense time	10.81 minutes		heads wt.=	405	lbs

total wt. 11,950 lbs (includes additional 15% on standpipe weight + Cyclone weight)

Heat Exchanger

Heat Exchanger Area (ft^2) 448 heat exchanger pipe thickness 0.25 in

volume of steel 9

total weight 4,560 lbs

total wt. **35,960** **lbs** weight for DSRP reactor system

COST

				>100,000 lb calc		
C.S. unit price for quantiy needed	2.260	$/lb	1990 $	1.928109822	2.26	
	2.415	$/lb	1996 $			<100,000 lb calc

Cost of installation $173,677

Total reactor cost **$477,612** includes cost of installation

Desulf and Regen transport reactor price calculation

DSRP

Regenerator Reactor

v (ft/sec) =	20		72000	ft/hr	air volume	139,951	cfh	HP-O2-N2
V (cfh) =	176,007				ROG volume	194,430	cfh	ROG
Area =	2.445	ft^2			regen sorbent flow	528,985	lb/hr	ZNS2RGEN
I.D.	21.614	in			regen sorbent vol.	8,816	cfh	
					sorbent vol%	5.01%		
thickness=	0.419							
shell wt.=	9,656	lbs			Corrosion depth	0.125	in	
heads wt.=	174	lbs			reactor height	100	ft	

total wt. 11,305 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

size vs. regen size 1

total wt. 11,305 lbs

Desulfurization Reactor

v (ft/sec) =	20		72000	ft/hr	coal gas in volume	2,531,530	cfh	RAW-CG
V (cfh) =	2,722,971				cg out volume	2,867,390	cfh	CG-CALC
Area =	37.82	ft^2			regen sorbent flow	1,410,630	lb/hr	ZNS
I.D.	85.01	in			regen sorbent vol.	23,511	cfh	
					sorbent vol%	0.86%		
thickness=	1.280348							
shell wt.=	116,135	lbs			Corrosion depth	0.125	in	
heads wt.=	8,227	lbs			reactor height	100	ft	

total wt. 143,017 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

size vs. desulf size 1

total wt. 143,017 lbs

total wt. 308,644 lbs weight for desulfurization and regeneration transport reactors

COST

				>100,000 lb calc	
C.S. unit price for quantiy needed	1.342	$/lb	1990 $	1.3421617	1.088
	1.434	$/lb	1996 $		<100,000 lb calc

Cost of installation $885,263

Total reactor cost $2,434,474 includes cost of installation

DSRP-500 Reactor Cost

DSRP

DSRP Reactor

v (ft/sec) gas =	3	10800 ft/hr	slipstream	76,726	cfh	SLIPSTREAM
V (cfh) =	236,095		ROG volume	153,894	cfh	ROG-COOL
space time -gas	33.33 seconds		reactor effluent	221,163	cfh	RXNPRD
v (ft/sec) cat =	2.3	8280 ft/hr	DSRP reactor Q	-31,370,000	BTU/hr	
			catalyst flow	612,229	lb/hr	
			catalyst vol.	10,204	cfh	
Area =	22.148 ft^2					
I.D.	65.058 in		catalyst vol%	5.56%		
thickness=	1.009 in					
shell wt.=	70,050 lbs		Corrosion depth	0.125	in	
heads wt.=	3,798 lbs		reactor height	100	ft	

total wt. 84,925 lbs (includes additional 15% for nozzles, manholes, etc.)

DSRP Standpipe

Cyclone (20% of reactor size)	16,985	Area =	22.15	ft^2
		I.D.	65.06	in
standpipe height	40 ft	thickness=	1.01	in
		shell wt.=	28,020	lbs
residense time	10.86 minutes	heads wt.=	3,798	lbs

total wt. 53,575 lbs (includes additional 15% on standpipe weight + Cyclone weight)

Heat Exchanger

Heat Exchanger Area (ft^2) 2174 heat exchanger pipe thickness 0.25 in

volume of steel 45

total weight 22,146 lbs

total wt. 160,646 lbs weight for DSRP reactor system

COST

			>100,000 lb calc	
C.S. unit price for quantiy needed	1.477 $/lb	1990 $	1.477409568	1.359
	1.579 $/lb	1996 $		<100,000 lb calc

Cost of installation $507,200

Total reactor cost $1,394,800 includes cost of installation

Appendix I
Sizing Reactors for the AHGP

Copies of the reactor system sizing calculations follow. They include estimates of the reactor system costs. The equations describe in *Appendix G - Calculation of Reactor Size* where used in the spreadsheet.

Desulf and Regen transport reactor price calculation

AHGP

N2 lift

v (ft/sec) =	50	180000 ft/hr	N2 in volume	42,761	cfh	N2SOURCE
V (cfh) =	64,033		N2 out volume	79,826	cfh	N2EXIT
Area =	0.356 ft^2		regen sorbent flow	164,358	lb/hr	FEO-ZNO
I.D.	8.245 in		regen sorbent vol.	2,739	cfh	
			sorbent vol%	4.28%		
thickness=	0.237					
shell wt.=	2,085 lbs		Corrosion depth	0.125	in	
heads wt.=	14 lbs		reactor height	100	ft	

key
caculated or constant values
inputed variables

total wt. 2,415 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

		residence time	10	min	
volume -sorbent	457 ft^3				
		heat removal	14,850,494	BTU/hr	RGENSTND
Heat Exchanger Area	1,029 ft^2	heat exchanger pipe thickness	0.25	in	
length of pipe	2,620 ft	heat exchanger pipe I.D.	1	in	
volume -heat exchanger	32 ft^3				
necessary standpipe volum	489 ft^3	standpipe height	60	ft	
Area =	8 ft^2				
I.D. =	39 in				
		Corrosion depth	0.125	in	
thickness=	0.650 in				
shell wt. =	16,085 lbs				
heads wt. =	863 lbs				

total wt. 19,491 lbs (includes additional 15% for nozzles, manholes,etc.

Desulfurization Reactor

			coal gas in volume	1,130,000	cfh	H2S-CG
v (ft/sec) =	20	72000 ft/hr	cg out volume	1,130,000	cfh	CLEAN-CG
V (cfh) =	1,135,506			164,358	lb/hr	COLDSORB
Area =	15.77 ft^2			166,009	lb/hr	STNDPIPE
I.D.	54.90 in		regen sorbent flow	330,367	lb/hr	
			regen sorbent vol.	5,506	cfh	
thickness=	0.871 in		sorbent vol%	0.48%		
shell wt.=	51,023 lbs		Corrosion depth	0.125	in	
heads wt.=	2,334 lbs		reactor height	100	ft	

total wt. 61,361 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

		residence time	1	min
volume -sorbent	2,767 ft^3			
Area =	28 ft^2	standpipe height	100	ft
I.D. =	71 in			
		Corrosion depth	0.125	in
thickness=	1.093 in			
shell wt. =	83,058 lbs			
heads wt. =	4,930 lbs			

total wt. 101,186 lbs (includes additional 15% for nozzles, manholes,etc.

Three Stage Regenerator

		number of reactors	2	
I.D. =	13.01 ft	standpipe height	45	ft
I.D. =	156 in			
thickness=	2.247 in			
shell wt. =	168,516 lbs	Corrosion depth	0.125	in
heads wt. =	48,735 lbs			

total wt. 260,701 lbs (includes additional 20% for cyclones, nozzles, manholes,etc.

total wt. 383,793 lbs weight for desulfurization and regeneration transport reactors

COST

			>100,000 lb calc
C.S. unit price for quantiy needed	1.303 $/lb	1990 $	1.30330926 1.01
	1.393 $/lb	1996 $	<100,000 lb calc

Cost of installation $1,068,941

Total reactor cost $2,939,588 includes cost of installation

SO2 Regenerator Sizing - Commercial Embodiment

Givens:	Revised (SO2 Regen) Case E-2	Calculated values:		Operating conditions/Gas Density Calc'ns:	
		Hold-up volume, ft3	2660.41667	Pressure, psig	275
		Diameter, ft	18.4047564	Pressure, psia	289.7
Sorbent circulation rate, lb/h	166010	X-section area, ft2	266.041667	MW of gas	64
Sorbent bulk density, lb/ft3	62.4	Calculated H/D	0.54333781	Bed Temp., C	600
Req'd rxtr residence time, hr	1	RG Vol. flow rate, acf/sec	21.8210028	Bed Temp., R	1571.67
Regen Gas v_{super}, cm/sec	2.5	RG flow rate, lb/hr	86366.3549	R, gas constant,	10.73
Desired H/D	2	Ratio of RG flow/sorbent, lb/lb	0.52024791	Gas density, lb/ft3	1.099429
		Calculated Bed Depth, ft			

Adjusted values:

Assumed Bed Depth, ft	10
SO2 needed ft3/hr	79812.5

Desulf and Regen transport reactor price calculation

AHGP

N2 lift

v (ft/sec) =	50	180000 ft/hr	N2 in volume	91,631	cfh	N2SOURCE
V (cfh) =	141,578		N2 out volume	175,069	cfh	N2EXIT
Area =	0.787 ft^2		regen sorbent flow	493,650	lb/hr	FEO-ZNO
I.D.	12.260 in		regen sorbent vol.	8,228	cfh	
			sorbent vol%	5.81%		
thickness=	0.292					
shell wt.=	3,815 lbs		Corrosion depth	0.125	in	
heads wt.=	39 lbs		reactor height	100	ft	

key
caculated or constant values
inputed variables

total wt. 4,432 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

		residence time	10	min
volume -sorbent	1371 ft^3			
		heat removal	49,966,040	BTU/hr RGENSTND
Heat Exchanger Area	3,462 ft^2	heat exchanger pipe thickness	0.25	in
length of pipe	8,817 ft	heat exchanger pipe I.D.	1	in
volume -heat exchanger	108 ft^3			
necessary standpipe volum	1,479 ft^3	standpipe height	60	ft
Area =	24.7 ft^2			
I.D. =	67.2 in			
		Corrosion depth	0.125	in
thickness=	1.039 in			
shell wt. =	44,713 lbs			
heads wt. =	4,176 lbs			

total wt. 56,222 lbs (includes additional 15% for nozzles, manholes,etc.

Desulfurization Reactor

			coal gas in volume	1,160,000	cfh	H2S-CG
v (ft/sec) =	20	72000 ft/hr	cg out volume	1,160,000	cfh	CLEAN-CG
V (cfh) =	1,176,539			493,650	lb/hr	COLDSORB
Area =	16.341 ft^2			498,670	lb/hr	STANDPIP
I.D.	55.881 in		regen sorbent flow	992,320	lb/hr	
			regen sorbent vol.	16,539	cfh	
thickness=	0.884 in		sorbent vol%	1.41%		
shell wt.=	52,734 lbs		Corrosion depth	0.125	in	
heads wt.=	2,456 lbs		reactor height	100	ft	

total wt. 63,468 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

		residence time	1	min
volume -sorbent	8,311 ft^3			
Area =	83 ft^2	standpipe height	100	ft
I.D. =	123 in			
		Corrosion depth	0.125	in
thickness=	1.803 in			
shell wt. =	237,425 lbs			
heads wt. =	24,424 lbs			

total wt. 301,127 lbs (includes additional 15% for nozzles, manholes,etc.

Three Stage Regenerators

		number of reactors	6	
I.D. =	12.99 ft	standpipe height	45	ft
I.D. =	156 in			
thickness=	2.243 in			
shell wt. =	167,848 lbs	Corrosion depth	0.125	in
heads wt. =	48,443 lbs			

total wt. 1,557,295 lbs (includes additional 20% for cyclones, nozzles, manholes,etc.

total wt. 1,919,076 lbs weight for desulfurization and regeneration transport reactors

COST

			>100,000 lb calc
C.S. unit price for quantiy needed	1.090 $/lb	1990 $	1.09039137 0.585
	1.165 $/lb	1996 $	<100,000 lb calc

Cost of installation $4,471,817

Total reactor cost $12,297,497 includes cost of installation

SO2 Regenerator Sizing - Commercial Embodiment

	AHGP-b (SO2 Regen)				
Givens:	**Case E-2**	**Calculated values:**		**Operating conditions/Gas Density Calc'ns:**	
		Hold-up volume, ft3	7948.71795	Pressure, psig	275
		Diameter, ft	31.8129385	Pressure, psia	289.7
Sorbent circulation rate, lb/h	496000	X-section area, ft2	794.871795	MW of gas	64
Sorbent bulk density, lb/ft3	62.4	Calculated H/D	0.31433751	Bed Temp., C	600
Req'd rxtr residence time, h	1	RG Vol. flow rate, acf/sec	65.1961774	Bed Temp., R	1571.67
Regen Gas v_{super}, cm/sec	2.5	RG flow rate, lb/hr	258042.961	R, gas constant,	10.73
Desired H/D	2	Ratio of RG flow/sorbent, lb/lb	0.52024791	Gas density, lb/ft3	1.099429
		Calculated Bed Depth, ft			
Adjusted values:					
Assumed Bed Depth, ft	10				
SO2 needed ft3/hr	238461.5385				

AHGP-c

Desulf and Regen transport reactor price calculation

AHGP

N2 lift

v (ft/sec) =	50	180000 ft/hr	N2 in volume	8,552	cfh	N2SOURCE	
V (cfh) =	13,240		N2 out volume	16,326	cfh	N2EXIT	
Area =	0.074 ft^2		regen sorbent flow	48,050	lb/hr	FEO-ZNO	
I.D.	3.749 in		regen sorbent vol.	801	cfh		key
			sorbent vol%	6.05%			caculated or constant values
thickness=	0.176						inputed variables
shell wt.=	704 lbs		Corrosion depth	0.125	in		
heads wt.=	2 lbs		reactor height	100	ft		

total wt. 812 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

		residence time	10	min	
volume -sorbent	133 ft^3				
		heat removal	48,050	BTU/hr	RGENSTND
Heat Exchanger Area	3.25 ft^2	heat exchanger pipe thickness	0.25	in	
length of pipe	8.27 ft	heat exchanger pipe I.D.	1	in	
volume -heat exchanger	0.10 ft^3				
necessary standpipe volume	134 ft^3	standpipe height	60	ft	
Area =	2.2 ft^2				
I.D. =	20.2 in				
		Corrosion depth	0.125	in	
thickness=	0.400 in				
shell wt. =	5,168 lbs				
heads wt. =	145 lbs				

total wt. 6,110 lbs (includes additional 15% for nozzles, manholes,etc.

Desulfurization Reactor

v (ft/sec) =	20	72000 ft/hr	coal gas in volume	1,120,000	cfh	H2S-CG
V (cfh) =	1,121,611		cg out volume	1,120,000	cfh	CLEAN-CG
				48,050	lb/hr	COLDSORB
Area =	15.58 ft^2			48,626	lb/hr	STANDPIP
I.D.	54.56 in		regen sorbent flow	96,676	lb/hr	
			regen sorbent vol.	1,611	cfh	
thickness=	0.867 in		sorbent vol%	0.14%		
shell wt.=	50,444 lbs		Corrosion depth	0.125	in	
heads wt.=	2,294 lbs		reactor height	100	ft	

total wt. 60,648 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

		residence time	1	min
volume -sorbent	810 ft^3			
Area =	8.10 ft^2	standpipe height	100	ft
I.D. =	38.55 in			
		Corrosion depth	0.125	in
thickness=	0.649 in			
shell wt. =	26,687 lbs			
heads wt. =	857 lbs			

total wt. 31,676 lbs (includes additional 15% for nozzles, manholes,etc.

Three Stage Regenerator

		number of reactors	1	
I.D. =	9.90 ft	standpipe height	45	ft
I.D. =	119 in			
thickness=	1.739 in			
shell wt. =	99,156 lbs	Corrosion depth	0.125	in
heads wt. =	21,807 lbs			

total wt. 145,156 lbs (includes additional 20% for cyclones, nozzles, manholes,etc.

total wt. 183,754 lbs weight for desulfurization and regeneration transport reactors

COST

			>100,000 lb calc
C.S. unit price for quantiy needed	1.447 $/lb	1990 $	1.44706713 1.298
	1.546 $/lb	1996 $	<100,000 lb calc

Cost of installation $568,244

Total reactor cost $1,562,672 includes cost of installation

SO2 Regenerator Sizing - Commercial Embodiment

AHGP-c (SO2 Regen)

Givens:	**Case E-2**	**Calculated values:**		**Operating conditions/Gas Density Calc'ns:**	
		Hold-up volume, ft3	769.23	Pressure, psig	275
		Diameter, ft	9.90	Pressure, psia	289.7
Sorbent circulation rate, lb/h	48000	X-section area, ft2	76.92	MW of gas	64
Sorbent bulk density, lb/ft3	62.4	Calculated H/D	1.01	Bed Temp., C	600
Req'd rxtr residence time, hr	1	RG Vol. flow rate, acf/sec	6.31	Bed Temp., R	1571.67
Regen Gas v_{super}, cm/sec	2.5	RG flow rate, lb/hr	24971.90	R, gas constant,	10.73
Desired H/D	2	Ratio of RG flow/sorbent, lb/lb	0.52	Gas density, lb/ft3	1.10
		Calculated Bed Depth, ft			

Adjusted values:

Assumed Bed Depth, ft	10
SO2 needed ft3/hr	23077

AHGP-100

Desulf and Regen transport reactor price calculation

(0.4211 the size of the AHGP case)

AHGP

N2 lift

v (ft/sec) =	50	180000 ft/hr	N2 in volume	18,007	cfh	N2SOURCE	
V (cfh) =	26,964		N2 out volume	33,615	cfh	N2EXIT	
Area =	0.150 ft^2		regen sorbent flow	69,211	lb/hr	FEO-ZNO	
I.D.	5.350 in		regen sorbent vol.	1,154	cfh		key
			sorbent vol%	4.28%			caculated or constant values
thickness=	0.198						inputed variables
shell wt.=	1,129 lbs		Corrosion depth	0.125	in		
heads wt.=	5 lbs		reactor height	100	ft		

total wt. 1,304 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

		residence time	10	min
volume -sorbent	192 ft^3			
		heat removal	6,253,543	BTU/hr RGENSTND
Heat Exchanger Area	433 ft^2	heat exchanger pipe thickness	0.25	in
length of pipe	1,103 ft	heat exchanger pipe I.D.	1	in
volume -heat exchanger	14 ft^3			
necessary standpipe volume	206 ft^3	standpipe height	60	ft
Area =	3.43 ft^2			
I.D. =	25.08 in			
		Corrosion depth	0.125	in
thickness=	0.466 in			
shell wt. =	7,478 lbs			
heads wt. =	260 lbs			

total wt. 8,899 lbs (includes additional 15% for nozzles, manholes,etc.

Desulfurization Reactor

v (ft/sec) =	20	72000 ft/hr	coal gas in volume	475,843	cfh	H2S-CG
V (cfh) =	478,162		cg out volume	475,843	cfh	CLEAN-CG
Area =	6.64 ft^2			69,211	lb/hr	COLDSORB
I.D.	35.62 in			69,906	lb/hr	STNDPIPE
			regen sorbent flow	139,118	lb/hr	
			regen sorbent vol.	2,319	cfh	
thickness=	0.609 in		sorbent vol%	0.48%		
shell wt.=	23,154 lbs		Corrosion depth	0.125	in	
heads wt.=	687 lbs		reactor height	100	ft	

total wt. 27,418 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

		residence time	1	min
volume -sorbent	1,165 ft^3			
Area =	11.7 ft^2	standpipe height	100	ft
I.D. =	46.2 in			
		Corrosion depth	0.125	in
thickness=	0.753 in			
shell wt. =	37,140 lbs			
heads wt. =	1,430 lbs			

total wt. 44,356 lbs (includes additional 15% for nozzles, manholes,etc.

Three Stage Regenerators

		number of reactors	1	
I.D. =	11.94 ft	standpipe height	45	ft
I.D. =	143 in			
thickness=	2.073 in			
shell wt. =	142,638 lbs	Corrosion depth	0.125	in
heads wt. =	37,858 lbs			

total wt. 216,595 lbs (includes additional 20% for cyclones, nozzles, manholes,etc.

total wt. 271,154 lbs weight for desulfurization and regeneration transport reactors

COST

			>100,000 lb calc
C.S. unit price for quantiy needed	1.367 $/lb	1990 $	1.36665221 1.137
	1.460 $/lb	1996 $	<100,000 lb calc

Cost of installation $791,924

Total reactor cost $2,177,791 includes cost of installation

SO2 Regenerator Sizing - Commercial Embodiment

	AHGP-100 (SO2 Regen)	Calculated values:		Operating conditions/Gas Density Calc'ns:	
Givens:	**Case E-2**	Hold-up volume, ft3	1,120.35	Pressure, psig	275
		Diameter, ft	11.94	Pressure, psia	289.7
Sorbent circulation rate, lb/h	69910	X-section area, ft2	112.04	MW of gas	64
Sorbent bulk density, lb/ft3	62.4	Calculated H/D	0.84	Bed Temp., C	600
Req'd rxtr residence time, hr	1	RG Vol. flow rate, acf/sec	9.19	Bed Temp., R	1571.67
Regen Gas v_{super}, cm/sec	2.5	RG flow rate, lb/hr	36,370.53	R, gas constant,	10.73
Desired H/D	2	Ratio of RG flow/sorbent, lb/lb	0.520	Gas density, lb/ft3	1.099429
		Calculated Bed Depth, ft			

Adjusted values:

Assumed Bed Depth, ft	10.00
SO2 needed ft3/hr	33611

AHGP-500

Desulf and Regen transport reactor price calculation (2.1055 the size of the AHGP case)

AHGP

N2 lift

v (ft/sec) =	50	180000 ft/hr		N2 in volume	90,033	cfh	N2SOURCE
V (cfh) =	134,821			N2 out volume	168,074	cfh	N2EXIT
Area =	0.75	ft^2		regen sorbent flow	346,056	lb/hr	FEO-ZNO
I.D.	11.96	in		regen sorbent vol.	5,768	cfh	
				sorbent vol%	4.28%		
thickness=	0.287592409						
shell wt.=	3,671	lbs		Corrosion depth	0.125	in	
heads wt.=	37	lbs		reactor height	100	ft	

key
- caculated or constant values
- inputed variables

total wt. 4,264 lbs (includes additional 15% for nozzles, manholes, etc.)

Regenerator Standpipe

			residence time	10	min
volume -sorbent	961	ft^3			
			heat removal	31,267,715	BTU/hr RGENSTND
Heat Exchanger Area	2,167	ft^2	heat exchanger pipe thickness	0.25	in
length of pipe	5,517	ft	heat exchanger pipe I.D.	1	in
volume -heat exchanger	68	ft^3			
necessary standpipe volume	1,029	ft^3	standpipe height	60	ft
Area =	17	ft^2			
I.D. =	56	in			
			Corrosion depth	0.125	in
thickness=	0.887	in			
shell wt. =	31,844	lbs			
heads wt. =	2,480	lbs			

total wt. 39,472 lbs (includes additional 15% for nozzles, manholes,etc.

Desulfurization Reactor

v (ft/sec) =	20	72000 ft/hr	coal gas in volume	2,379,215	cfh	H2S-CG
V (cfh) =	2,390,808		cg out volume	2,379,215	cfh	CLEAN-CG
Area =	33.21	ft^2		346,056	lb/hr	COLDSORB
I.D.	79.66	in		349,532	lb/hr	STNDPIPE
			regen sorbent flow	695,588	lb/hr	
			regen sorbent vol.	11,593	cfh	
thickness=	1.208	in	sorbent vol%	0.48%		
shell wt.=	102,638	lbs	Corrosion depth	0.125	in	
heads wt.=	6,813	lbs	reactor height	100	ft	

total wt. 125,868 lbs (includes additional 15% for nozzles, manholes, etc.)

Desulfurization Standpipe

			residence time	1	min
volume -sorbent	5,826	ft^3			
Area =	58	ft^2	standpipe height	100	ft
I.D. =	103	in			
			Corrosion depth	0.125	in
thickness=	1.530	in			
shell wt. =	168,662	lbs			
heads wt. =	14,526	lbs			

total wt. 210,666 lbs (includes additional 15% for nozzles, manholes,etc.

Three Stage Regenerators

			number of reactors	5	
I.D. =	11.94	ft	standpipe height	45	ft
I.D. =	143	in			
thickness=	2.073	in			
shell wt. =	142,618	lbs	Corrosion depth	0.125	in
heads wt. =	37,850	lbs			

total wt. 1,082,809 lbs (includes additional 20% for cyclones, nozzles, manholes,etc.

total wt. 1,337,211 lbs weight for desulfurization and regeneration transport reactors

COST

			>100,000 lb calc	
C.S. unit price for quantiy needed	1.129 $/lb	1990 $	1.12859015	0.661
	1.206 $/lb	1996 $		<100,000 lb calc

Cost of installation $3,225,118

Total reactor cost **$8,869,074** includes cost of installation

SO2 Regenerator Sizing - Commercial Embodiment

Givens:	AHGP-500 (SO2 Regen) Case E-2	Calculated values:		Operating conditions/Gas Density Calc'ns:	
		Hold-up volume, ft3	5,600.96	Pressure, psig	275
		Diameter, ft	26.70	Pressure, psia	289.7
Sorbent circulation rate, lb/h	349500	X-section area, ft2	560.10	MW of gas	64
Sorbent bulk density, lb/ft3	62.4	Calculated H/D	0.37	Bed Temp., C	600
Req'd rxtr residence time, hr	1	RG Vol. flow rate, acf/sec	45.94	Bed Temp., R	1571.67
Regen Gas v_{super}, cm/sec	2.5	RG flow rate, lb/hr	181,826.64	R, gas constant,	10.73
Desired H/D	2	Ratio of RG flow/sorbent, lb/lb	0.520	Gas density, lb/ft3	1.099429
		Calculated Bed Depth, ft			

Adjusted values:

Assumed Bed Depth, ft	10.00
SO2 needed ft3/hr	168,029

Appendix J
Power Generation Achievable from Clean Coal Gas

Two sources where used in determining the power generated by the clean coal gas. The Sierra power generating facility was used as the basis for determining the power generating capacity coal gas.

Sierra Clean Coal Gas Feed

H_2 (lbmole/hr)	CO (lbmole/hr)	Power Generation (MW)
5760	7570	260

The individual contribution of the H_2 and CO where determined assuming there relative contribution was consistent with their standard heats of combustion.

Standard heat of combustion (Felled & Rousseau):
$\Delta H^o_{comb}\,(H_2) = -3.605E\text{-}2$ MW hr/ lbmole $\Delta H^o_{comb}\,(CO) = -3.569E\text{-}2$ MW hr/ lbmole

Power generation can be expressed:

$$E\,[5760\,\Delta H_C\,(H_2) + 7570\,\Delta H_C\,(CO)] = 260\text{ MW}$$

where:

E = Efficiency of power generation

assuming:

$\Delta H_C\,(CO) = 0.99\,\Delta H_C\,(H_2)$

and substituting gives:

$$13{,}254\,E\,\Delta H_C(H_2) = 260\text{ MW}$$

$$E\,\Delta H_C(H_2) = 0.0196\text{ MW hr / lbmole}$$

therefore

$$E\,\Delta H_C(CO) = 0.0194\text{ MW hr / lbmole}$$

The values calculated above can be used to write a power generation expression.

Power Generation {MW} = 0.0196 (H_2 {lbmoles/hr}) + 0.0194 (CO {lbmoles/hr})

The plants power generation is determined by inserting the clean coal gas flows for H_2 and CO into the above equation. HGD coal gas consumption is assessed as a debit equivalent to the cost of the lost power generation. The power generation lost is determined by inserting the difference in the dirty coal gas and clean coal gas molar flow rates into the above equation. The cost of the electricity is taken as $0.04 per kWh. The plant has been assumed to be in operation 90% of the year.

Summary of Power Generation Calculations

simulation	H_2 clean	H_2 in	CO clean	CO in	MW made	MW lost
DSRP	11,444.58	11,765.37	212,200.52	218,162.00	258.25	7.248
DSRP-b	11,450.19	12,468.32	212,276.67	231,196.50	258.35	23.003
DSRP-c	11,443.82	11,535.37	212,195.77	213,897.17	258.24	2.069
DSRP-100	4,819.31	4,954.40	89,357.59	91,868.05	108.75	3.052
DSRP-500	24,110.94	24,772.09	447,055.34	459,341.97	544.06	14.938
AHGP	11,355.75	11,510.68	213,439.25	213,439.25	258.24	1.506
AHGP-b	11,175.21	11,646.28	215,953.67	215,953.67	258.23	4.580
AHGP-c	11,419.66	11,464.48	212,582.61	212,582.61	258.27	0.436
AHGP-100	4,781.91	4,847.15	89,879.27	89,879.27	108.74	0.634
AHGP-500	23,909.53	24,235.73	449,396.34	449,396.34	543.72	3.172

Appendix K
Calculation of Reactor Pressure Drops

Pressure drops for transport reactors have been calculated assuming the pressure drops are related to the energy required to lift the sorbent / catalyst to the top of the reactor.

Energy balance for lifting solid to top of reactor:

$$\Delta E_{PART} = \Delta E_{GAS}$$

$$m_{PART}\ (g / g_C)\ h = \Delta P\ m_{GAS} / \rho_{GAS}$$

$$\Delta P = m_{PART}\ (g / g_C)\ h\ \rho_{GAS}\ / m_{GAS}$$

$$\Delta P_{REACTOR} = 1.5 \text{ (Energy to lift particle)}$$

<u>DSRP Regeneration Reactor</u>

$$\Delta P = 1.5\ m_{PART}\ (g / g_C)\ h\ \rho_{GAS}\ / m_{GAS}$$

m_{PART} = sorbent mass flow, ZNS2RGEN & ZNO average

$(g / g_C) = 1\ lb_f/lb_m$

h = reactor height, defined in Appendix H

ρ_{GAS} = gas density, HP-O2-N2 & ROG average

m_{GAS} = gas mass flow, HP-O2-N2 & ROG average

<u>DSRP Regeneration Reactor (DSRP)</u>

$$\Delta P = 1.5\ (250{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.5\ lb_m/ft^3) / (40{,}000\ lb_m/hr)\ (\ 1\ ft^3 / 144\ in^2)$$

$$\Delta P = 3.32 \text{ psi}$$

<u>DSRP Regeneration Reactor (DSRP-b)</u>

$$\Delta P = 1.5\ (1{,}200{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.5\ lb_m/ft^3) / (130{,}000\ lb_m/hr)(\ 1\ ft^3 / 144\ in^2)$$

$$\Delta P = 4.8 \text{ psi}$$

<u>DSRP Regeneration Reactor (DSRP-c)</u>

$$\Delta P = 1.5\ (71{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.5\ lb_m/ft^3) / (12{,}000\ lb_m/hr)(\ 1\ ft^3 / 144\ in^2)$$

$$\Delta P = 3.2 \text{ psi}$$

<u>DSRP Regeneration Reactor (DSRP-100) (DSRP-500)</u>

same as base case $\Delta P = 3.3$ psi

DSRP Reactor

$\Delta P = 1.5\ m_{PART}\ (g / g_C)\ h\ \rho_{GAS}\ /\ m_{GAS}$

m_{PART} = catalyst mass flow, Appendix H

$(g / g_C) = 1\ lb_f/lb_m$

h = reactor height, defined in Appendix H

ρ_{GAS} = gas density, ROG-COOL & RXNPRD average

m_{GAS} = gas mass flow, RXNPRD

DSRP Reactor (DSRP)

$\Delta P = 1.5\ (300{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.53\ lb_m/ft^3)\ /\ (55{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 3.0$ psi

DSRP Reactor (DSRP-b)

$\Delta P = 1.5\ (1{,}000{,}000\ lb_m/hr)(1\ lb_f/lb_m)\ (100\ ft)\ (0.53\ lb_m/ft^3)\ /\ (185{,}000\ lb_m/hr)(1\ ft^3/ 144\ in^2)$

$\Delta P = 3.0$ psi

DSRP Reactor (DSRP-c)

$\Delta P = 1.5\ (79{,}000\ lb_m/hr)(1\ lb_f/lb_m)\ (100\ ft)\ (0.55\ lb_m/ft^3)\ /\ (15{,}000\ lb_m/hr)(1\ ft^3/ 144\ in^2)$

$\Delta P = 3.0$ psi

DSRP Reactor (DSRP-100) (DSRP-500)

same as base case $\Delta P = 3.0$ psi

DSRP Desulfurization Reactor

$\Delta P = 1.5\ m_{PART}\ (g / g_C)\ h\ \rho_{GAS}\ /\ m_{GAS}$

m_{PART} = sorbent mass flow, ZNS

$(g / g_C) = 1\ lb_f/lb_m$

h = reactor height, defined in Appendix H

ρ_{GAS} = gas density, RAW-CG & CG-CALC average

m_{GAS} = gas mass flow, CG-CALC

DSRP Desulfurization Reactor (DSRP)

$\Delta P = 1.5\ (670{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (510{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 0.6\ psi$

DSRP Desulfurization Reactor (DSRP-b)

$\Delta P = 1.5\ (2{,}700{,}000\ lb_m/hr)(1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (660{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 1.6\ psi$

DSRP Desulfurization Reactor (DSRP-c)

$\Delta P = 1.5\ (72{,}000\ lb_m/hr)(1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (460{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 0.06\ psi$

DSRP Desulfurization Reactor (DSRP-100) (DSRP-500)

same as base case $\Delta P = 0.6\ psi$

AHGP Desulfurization Reactor

$\Delta P = 1.5\ m_{PART}\ (g / g_C)\ h\ \rho_{GAS} / m_{GAS}$

m_{PART} = sorbent mass flow, STNDPIPE + COLDSORB

$(g / g_C) = 1\ lb_f/lb_m$

h = reactor height, defined in Appendix I

ρ_{GAS} = gas density, H2S-CG & CLEAN-CG average

m_{GAS} = gas mass flow, CLEAN-CG

AHGP Desulfurization Reactor (AHGP-100 and AHGP-500 results will be consistent)

$\Delta P = 1.5\ (330{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (450{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 0.3\ psi$

AHGP-b Desulfurization Reactor

$\Delta P = 1.5\ (990{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (460{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 0.9\ psi$

AHGP-c Desulfurization Reactor

$\Delta P = 1.5\ (97{,}000\ lb_m/hr)\ (1\ lb_f/lb_m)\ (100\ ft)\ (0.4\ lb_m/ft^3) / (440{,}000\ lb_m/hr)\ (1\ ft^3 / 144\ in^2)$

$\Delta P = 0.09\ psi$

The pressure drop through the bubble bed regenerator is calculated as the sum of the static head in each stage times 1.3.

AHGP 3-Stage Regenerator Reactor

$$\Delta P = 1.3\ g/g_C\ (\rho\ h_{top\text{-}stage} + \rho\ h_{stage2} + \rho\ h_{bottom\text{-}stage})\ (1/144)$$

m_{PART} = sorbent mass flow, FES-ZNS

$(g / g_C) = 1\ lb_f/lb_m$

h = reactor stage height,

ρ_{GAS} = average of density of streams entering and exiting the reactor stage

AHGP 3-Stage Regenerator Reactors

$$\Delta P = 1.3\ (1\ lb_f/lb_m)\ [\ (3.66\ lb_m/ft^3)\ (5.0\ ft) + (3.20\ lb_m/ft^3)\ (10\ ft) + (3.40\ lb_m/ft^3)\ (2.5\ ft)]\ (\ 1\ ft^3\ /\ 144\ in^2)$$

$$\Delta P = 0.5\ psi$$

Appendix L
Summary of the Process Pressure Drops

This appendix contains lists of the calculated pressure drops for the DSRP and AHGP at the various feed conditions.

DSRP pressure drops are used to determine the pressure rise needed from the RECYCOMP (sends tailgas to the Desulfurization reactor) and PRESAIR (pressurizes the air fed to the regenerator)   Reactor pressure drops are calculated in Appendix H. Pressure drops in other equipment has been assigned without calculations.

Having streams enter the DSRP Reactor at the same pressure (bold pressures) was the starting point for the calculations.

DSRP (base case) & DSRP-100 & DSRP-500

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
PRESAIR	13.7 psia inlet P	278.9
pipe [P-02-N2]	0	278.9
AIR-HX (shell)	2.0	276.9
pipe [HP-O2-N2]	0	276.9
REGENERATOR	3.3	273.6
pipe [ROG]	0	273.6
AIR-HX (tube)	2.0	271.6
pipe [ROG-COOL]	0	**271.6**
DSRP	2.0	268.6
pipe [RXNPRD]	0	268.6
PD-COOLR	2.0	266.6
pipe [COOLPRD]	0	266.6
High Press. Cond.	2.0	264.6
pipe [TAILGAS]	0	264.6
VALVE	2.6	262.0
pipe [TAILGAS2]	0	262.0
RECYCOMP		275

Coal Gas Slipstream Pressure

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
Desulfurization Reactor	0.6	274.4
pipe [SLIPSTRM]	0	274.4
VALVE2	2.8	271.6
pipe [SLPSTRM]	0	**271.6**

DSRP-b

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
PRESAIR	13.7 psia inlet P	279.4
pipe [P-O2-N2]	0	279.4
AIR-HX (shell)	2.0	277.4
pipe [HP-O2-N2]	0	277.4
REGENERATOR	4.8	272.6
pipe [ROG]	0	272.6
AIR-HX (tube)	2.0	270.6
pipe [ROG-COOL]	0	**270.6**
DSRP	3.0	267.6
pipe [RXNPRD]	0	267.6
PD-COOLR	2.0	265.6
pipe [COOLPRD]	0	265.6
High Press. Cond.	2.0	263.6
pipe [TAILGAS]	0	263.6
VALVE	2.6	261.0
pipe [TAILGAS2]	0	261.0
RECYCOMP		275

Coal Gas Slipstream Pressure

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
Desulfurization Reactor	1.6	273.4
pipe [SLIPSTRM]	0	273.4
VALVE2	2.8	270.6
pipe [SLPSTRM]	0	**270.6**

DSRP-c

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
PRESAIR	13.7 psia inlet P	279.3
pipe [P-O2-N2]	0	279.3
AIR-HX (shell)	2.0	277.3
pipe [HP-O2-N2]	0	277.3
REGENERATOR	3.2	274.1
pipe [ROG]	0	274.1
AIR-HX (tube)	2.0	272.1
pipe [ROG-COOL]	0	**272.1**
DSRP	3.0	269.1
pipe [RXNPRD]	0	269.1
PD-COOLR	2.0	267.1
pipe [COOLPRD]	0	267.1
High Press. Cond.	2.0	265.1
pipe [TAILGAS]	0	265.1
VALVE	2.6	262.5
pipe [TAILGAS2]	0	262.5
RECYCOMP		275

DSRP-c

Coal Gas Slipstream Pressure

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
Desulfurization Reactor	0.06	274.9
pipe [SLIPSTRM]	0	274.9
VALVE2	2.8	272.1
pipe [SLPSTRM]	0	**272.1**

AHGP pressure drop calculations determine the required ΔP for the SO2-COMP, compressor. The pressure drop balance is done to insure the SO_2 loop with maintain desired pressure. The set pressure (bold) in the SO_2 loop is the pressure at the 3-Stage Regenerator exit. This pressure is set to equal the calculated exit pressure of the AHGP Desulfurization reactor (Appendix K).

AHGP (base case), & AHGP-100 & AHGP-500

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
3-Stage Regenerator	0.5 (Append. K)	**274.7**
pipe [COOLS2]	0	274.7
HEATX (tube)	2.0	272.7
pipe [S2V+L]	0	272.7
COND-EQ	2.0	270.7
pipe [IN-COND]	0	270.7
DEMISTR	5	265.7
pipe [UNP-RSO2]	0	265.7
SO2-COMP		279.2
pipe [RCYHEATR]	0	279.2
RCYHEATR	2.0	277.2
pipe [WARMRCY]	0	277.2
HEATX (shell)	2.0	275.2
pipe [FEEDRG1]	0	275.2 to 3-Stage Regenerator

AHGP-b

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
3-Stage Regenerator	0.5 (Append. K)	**274.1**
pipe [COOLS2]	0	274.1
HEATX (tube)	2.0	272.1
pipe [S2V+L]	0	272.1
COND-EQ	2.0	270.1
pipe [IN-COND]	0	270.1
DEMISTR	5	265.1
pipe [UNP-RSO2]	0	265.1
SO2-COMP		278.6
pipe [RCYHEATR]	0	278.6
RCYHEATR	2.0	276.6
pipe [WARMRCY]	0	276.6
HEATX (shell)	2.0	274.6
pipe [FEEDRG1]	0	274.6 to 3-Stage Regenerator

AHGP-c

Equipment	ΔP drop (psi)	P_{EXIT} (psia)
3-Stage Regenerator	0.5 (Append. K)	**274.9**
pipe [COOLS2]	0	274.9
HEATX (tube)	2.0	272.9
pipe [S2V+L]	0	272.9
COND-EQ	2.0	270.9
pipe [IN-COND]	0	270.9
DEMISTR	5	265.9
pipe [UNP-RSO2]	0	265.9
SO2-COMP		279.4
pipe [RCYHEATR]	0	279.4
RCYHEATR	2.0	277.4
pipe [WARMRCY]	0	277.4
HEATX (shell)	2.0	275.4
pipe [FEEDRG1]	0	275.4 to 3-Stage Regenerator

Appendix M
Summary of Major HGD Equipment

The following tables list equipment required for both HGD processes under various feed conditions. Equipment specifications are also listed in the tables.

DSRP - base Process Equipment Specifications

	DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500
REACTORS					
Desulfurization reactor					
height (ft)	100	100	100	100	100
diameter (ft)	4.9	5.6	4.6	3.2	7.1
weight (lbs)	70,000	90,000	63,000	31,000	140,000
Desulf. standpipe					
height (ft)	100	100	100	100	100
diameter (ft)	4.9	5.6	4.6	3.2	7.1
weight (lbs)	70,000	90,000	63,000	31,000	140,000
Regeneration reactor					
height (ft)	100	100	100	100	100
diameter (ft)	1.3	2.3	0.66	0.82	1.8
weight (lbs)	6,000	17,000	2,000	3,000	11,000
Regen. standpipe					
height (ft)	100	100	100	100	100
diameter (ft)	1.3	2.3	0.66	0.82	1.8
weight (lbs)	6,000	17,000	2,000	3,000	11,000
DSRP Reactor					
height (ft)	100	100	100	100	100
diameter (ft)	3.8	7.0	2.0	2.5	5.4
weight (lbs)	43,000	140,000	13,000	19,000	85,000
DSRP standpipe					
height (ft)	40	40	40	40	40
diameter (ft)	3.8	7.0	2.0	2.5	5.4
weight (lbs)	27,000	89,000	8,000	12,000	540,000
COMPRESSORS					
PRESAIR					
acfh	570,000	1,800,000	160,000	240,000	1,200,000
Pin (psia)	13.7	13.7	13.7	13.7	13.7
Pout (psia)	278.9	279.4	279.3	278.9	278.9
power (hp)	3,300	10,000	900	1,400	6,900
stages	6	6	6	6	6
RECOMP					
acfh	49,000	170,000	14,000	21,000	100,000
Pin (psia)	264.4	261	262.5	264.4	264.4
Pout (psia)	275	275	275	275	275
power (hp)	59	227	17	25	124
stages	1	1	1	1	1
HEAT EXCHANGERS					
AIRHX					
Duty (BTU/hr)	4,300,000	14,000,000	1,200,000	1,900,000	9,600,000
Area (ft^2)	700	2,200	200	300	1,400
tube mat.	SS 310	SS 310	SS 310	SS 310	SS 310
shell mat.	SS 304	SS 304	SS 304	SS 304	SS 304
DSRP					
Duty (BTU/hr)	15,000,000	51,000,000	4,000,000	6,500,000	31,000,000
Area (ft^2)	1,000	3,600	280	450	2,200
tube mat.	SS 310	SS 310	SS 310	SS 310	SS 310
PDCOOLR					
Duty (BTU/hr)	5,200,000	17,000,000	1,400,000	2,200,000	11,000,000
Area (ft^2)	1,000	3,200	300	300	2,000
tube mat.	SS 310	SS 310	SS 310	SS 310	SS 310
shell mat.	SS 310	SS 310	SS 310	SS 310	SS 310
MISC.					
High Pressure Condenser					
Duty (BTU/hr)	10,500,000	35,100,000	2,940,000	4,320,000	21,600,000
Material	SS 310	SS 310	SS 310	SS 310	SS 310
VAPORIZR					
Duty (BTU/hr)	550,000	1,900,000	150,000	230,000	1,100,000
Material	SS 310	SS 310	SS 310	SS 310	SS 310
Storage Tank					
vol. (ft^3)	5,600	18,000	1,600	2,400	11,000
Material	SS 310	SS 310	SS 310	SS 310	SS 310

AHGP Process Equipment Specifications

		AHGP	AHGP - b	AHGP - c	AHGP - 100	AHGP -500
REACTORS						
Desulfurization reactor						
	height (ft)	100	100	100	100	100
	diameter (ft)	4.58	4.66	4.55	2.97	6.64
	weight (lbs)	61,361	63,000	61,000	27,000	130,000
Desulf. standpipe						
	height (ft)	100	100	100	100	100
	diameter (ft)	5.92	10.25	3.21	3.85	8.58
	weight (lbs)	100,000	300,000	32,000	44,000	210,000
Regeneration reactor						
	# of reactors	2	6	1	1	5
	height (ft)	45	45	45	45	45
	diameter (ft)	13.0	13.0	0.8	11.9	11.9
	weight (lbs)	260,000	1,600,000	150,000	270,000	1,000,000
Regen. standpipe & RGENSTAND						
	height (ft)	60	60	60	60	60
	diameter (ft)	3.25	5.6	1.68	2.1	4.7
	weight (lbs)	19,000	56,000	6,100	8,900	39,000
	Duty (BTU/hr)	15,000,000	50,000,000	48,000	6,300,000	31,000,000
N2 Lift						
	height (ft)	100	100	100	100	100
	diameter (ft)	0.69	1.02	0.31	0.45	1.00
	weight (lbs)	2,400	4,400	800	1,300	4,300
COMPRESSORS						
CON-COMP						
	acfh	1,500	4,400	400	600	3,200
	Pin (psia)	15	15	15	15	15
	Pout (psia)	279	279	279	279	279
	power (hp)	8	26	2	3	17
	stages	1	1	1	1	1
LIFTCOMP						
	acfh	43,000	92,000	8,600	18,000	91,000
	Pin (psia)	272	272	272	272	272
	Pout (psia)	275	275	275	275	275
	power (hp)	13	28	3	5	27
	stages	1	1	1	1	1
SO2-COMP						
	acfh	29,000	85,000	8,400	12,000	61,000
	Pin (psia)	266	265	266	266	266
	Pout (psia)	279	279	279	279	279
	power (hp)	38	114	11	16	80
	stages	1	1	1	1	1
HEAT EXCHANGERS						
N2-COOLR						
	Duty (BTU/hr)	7,020,000	15,800,000	1,480,000	3,130,000	15,700,000
	Area (ft^2)	1,100	2,600	210	470	2,300
	tube mat.	SS 304	SS 304	SS 304	SS 304	SS 304
	shell mat.	SS 304	SS 304	SS 304	SS 304	SS 304
HEATX						
	Duty (BTU/hr)	5,100,000	15,000,000	1,500,000	2,100,000	11,000,000
	Area (ft^2)	1,600	3,600	500	700	3,500
	tube mat.	SS 310	SS 310	SS 310	SS 310	SS 310
	shell mat.	SS 310	SS 310	SS 310	SS 310	SS 310
RCYHEATR						
	Duty (BTU/hr)	2,530,000	6,070,000	697,000	1,070,000	5,330,000
	Area (ft^2)	3,200	7,800	570	1,300	6,700
	tube mat.	SS 310	SS 310	SS 310	SS 310	SS 310
	shell mat.	SS 310	SS 310	SS 310	SS 310	SS 310
MISC.						
COND-EQ						
	Duty (BTU/hr)	5,380,000	16,000,000	1,560,000	2,400,000	12,000,000
	Material	SS 310	SS 310	SS 310	SS 310	SS 310
DEMISTR						
	Duty (BTU/hr)	0	0	0	0	0
	Material	SS 310	SS 310	SS 310	SS 310	SS 310
LP-COND						
	vol. (ft^3)	30	100	10	10	70
	Material	SS 310	SS 310	SS 310	SS 310	SS 310
Storage Tank						
	vol. (ft^3)	5,600	18,000	1,600	2,400	11,000
	Material	SS 310	SS 310	SS 310	SS 310	SS 310

Appendix N
Summary of HGD Costs

The following pages are taken from an Excel spreadsheet containing the culmination of all costs and benefits for all simulated Hot Gas Desulfurization processes.

DSRP costs

Equipment -Sulfur side

Type	unit	DSRP Price	DSRP-b Price	DSRP-c Price	DSRP-100 Price	DSRP-500 Price	Mat. of Construction	Purchase date	Purchase price ref.	date of calculation
Heat Exchangers										
	AIRHX	$33,500	$71,500	$17,900	$19,400	$55,300	SS304 / SS310 tubes	June, 1996	aspen DAIRHX	1/22/98
	PDCOOLR	$63,400	$126,600	$25,200	$42,000	$90,400	SS310 (calc w SS316)	June, 1996	aspen	1/22/98
Tanks										
7 days	Sulfur Storage	$125,500	$205,400	$65,000	$80,000	$171,000	SS310 (calc w SS316)	June, 1996	aspen	11/6/97
Condenser										
	High Pressure	$40,400	$82,200	$18,500	$21,900	$59,600	SS310 (calc w SS316)	June, 1996	aspen	
Vaporiser										
	VAPORIZR	$16,100	$17,800	$15,900	$15,200	$16,700	SS310 (calc w SS316)	June, 1996	aspen	
Compressor										
	RECOMP	$52,900	$52,900	$52,900	$52,900	$52,900	Carbon Steel	June, 1996	aspen	
	PRESAIR	$844,000	$2,680,000	$241,000	$416,000	$1,740,000		1997 Ingesoll-Rand Centac Pricing		10/20/97
Reactors										
	Desulf & Regen	$1,328,000	$1,772,000	$1,169,000	$728,000	$2,434,000	SS310 (calc w SS316)	June, 1996 (w install)	P&T calc	10/7/97
	DSRP reactor	$812,129	$2,134,355	$367,075	$477,612	$1,394,800	SS310 (calc w SS316)	June, 1996 (w install)	P&T calc	10/16/97
Pipes										
	pipe lines									
	totals	$3,315,929	$7,142,755	$1,972,475	$1,853,012	$6,014,700				

Equipment -Steam side

Type	unit	DSRP Price	DSRP-b Price	DSRP-c Price	DSRP-100 Price	DSRP-500 Price	Mat. of Construction	date		Purchase price ref.
Heat Exchangers										
	LCOOLR	$7,600	$8,100	$0	$6,800	$7,600		June, 1996	aspen	11/26/97
	VCOOLR	$7,000	$8,400	$6,700	$6,800	$7,600		June, 1996	aspen	11/26/97
Pumps										
	PTOWR	$4,200	$8,000	$2,800	$3,200	$5,500		June, 1996	aspen	11/26/97
	PHOTH2O	$1,000	$3,500	$0	$400	$3,500		price quote from General Pumps		
	PSTEAM	$57,400	$75,100	$57,400	$57,400	$59,300		June, 1996	aspen	11/26/97
	totals	$77,200	$103,100	$66,900	$74,600	$83,500				

Expendatures

		DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500		cost ref.	
Electrical									
Pumps & Compressors									
kW	RECYCOMP	59	227	5	7	37	ASPEN generated power requirements		
kW	PRESAIR	3282	10414	900	999	4889	ASPEN generated power requirements		
kW	Steam pumps	76	193	30	32	160	ASPEN steam simulations		11/26/97
Light & instruments									
kW	misc.	683	683	683	683	683	20% base case pump & compressor requirements		
TOTAL	kW	4100.4	11517.4	1618.4	1721	5769			
	unit cost $/kWh	0.04	0.04	0.04	0.04	0.04	Self-gen. (Jan. 1990)	Peters & Timmeraus	
90 % op	Cost $/yr	$1,293,988	$3,634,615	$510,728	$543,234	$1,820,690			
Cooling Water									
	lbs/hr	149,000	500,000	25,000	62,744	313,720	ASPEN Complete Steam Generation Scheme simulations		
	unit value $/lb	2.6E-05	2.6E-05	2.6E-05	2.6E-05	2.6E-05	Tower (Jan. 1990)	Peters & Timmeraus	
90 % op	Cost $/yr	$21,854	$73,336	$3,667	$9,203	$46,014			
Oxygen									
	lbs/hr	0	0	0	0	0			
	unit value $/lb								
	Cost $/yr								
Addtional Employees									
	Engineers	2	2	2	2	2			
	unit cost	$100,000	$100,000	$100,000	$100,000	$100,000			
	Maintenance	2	2	2	2	2			
	unit cost	$70,000	$70,000	$70,000	$70,000	$70,000			
	Cost $/yr	$340,000	$340,000	$340,000	$340,000	$340,000			
Consumed Coal Gas									
	MW lost	7	23	2	3	15	Appendix J		
	unit cost $/MWh	40	40	40	40	40	Self-gen. (Jan. 1990)	Peters & Timmeraus	
	Cost $/yr	$2,287,295	$7,259,195	$652,927	$963,138	$4,714,074			
	totals (yearly)	$3,943,137	$11,307,146	$1,507,322	$1,855,574	$6,920,778			

Benefits

		DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500	Condition	value ref.	date of calc.
Sulfur Recovered									
	lbs/hr	5,840	18,590	1,667	2,460	12,300			
90% op	tons/year	23,037	73,332	6,576	9,704	48,520			11/4/97
	unit value $/ton	50	50	50	50	50	low purity	Chem. Eng. Progress 1996	
	Revenue $/yr	$1,151,852	$3,666,599	$328,791	$485,198	$2,425,991			
Steam Generation									
	lbs/hr	23,200	77,700	6,160	9,800	48,800	950 psia, 441 C		11/4/97
	unit value $/lb	0.0039	0.0039	0.0039	0.0039	0.0039	500 psig, (Jan. 1990)	Peters and Timmeraus	
90% op	Revenue $/yr.	$713,833	$2,390,725	$189,535	$301,533	$1,501,511			
	totals (yearly)	$1,865,685	$6,057,324	$518,326	$786,731	$3,927,501			

	DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500
YEARLY COST	$2,077,452	$5,249,823	$988,996	$1,068,843	$2,993,277
EQUIPMENT COSTS	$3,393,129	$7,245,855	$2,039,375	$1,927,612	$6,098,200

DSRP costs

Equipment -Sulfur side

Type	unit	DSRP Price	DSRP-b Price	DSRP-c Price	DSRP-100 Price	DSRP-500 Price	Mat. of Construction	Purchase date	Purchase price ref.	date of calculation
Heat Exchangers										
	AIRHX	$33,500	$71,500	$17,900	$19,400	$55,300	SS304 / SS310 tubes	June, 1996	aspen DAIRHX	1/22/98
	PDCOOLR	$63,400	$126,600	$25,200	$42,000	$90,400	SS310 (calc w SS316)	June, 1996	aspen	1/22/98
Tanks										
7 days	Sulfur Storage	$125,500	$205,400	$65,000	$80,000	$171,000	SS310 (calc w SS316)	June, 1996	aspen	11/6/97
Condenser										
	High Pressure	$40,400	$82,200	$18,500	$21,900	$59,600	SS310 (calc w SS316)	June, 1996	aspen	
Vaporiser										
	VAPORIZR	$16,100	$17,800	$15,900	$15,200	$16,700	SS310 (calc w SS316)	June, 1996	aspen	
Compressor										
	RECOMP	$52,900	$52,900	$52,900	$52,900	$52,900	Carbon Steel	June, 1996	aspen	
	PRESAIR	$844,000	$2,680,000	$241,000	$416,000	$1,740,000		1997 Ingesoll-Rand Centac Pricing		10/20/97
Reactors										
	Desulf & Regen	$1,328,000	$1,772,000	$1,169,000	$728,000	$2,434,000	SS310 (calc w SS316)	June, 1996 (w install)	P&T calc	10/7/97
	DSRP reactor	$812,129	$2,134,355	$367,075	$477,612	$1,394,800	SS310 (calc w SS316)	June, 1996 (w install)	P&T calc	10/16/97
Pipes										
	pipe lines									
	totals	$3,315,929	$7,142,755	$1,972,475	$1,853,012	$6,014,700				

DSRP costs

Equipment -Steam side

Type	unit	DSRP Price	DSRP-b Price	DSRP-c Price	DSRP-100 Price	DSRP-500 Price	Mat. of Construction	date		Purchase price ref.
Heat Exchangers										
	LCOOLR	$7,600	$8,100	$0	$6,800	$7,600		June, 1996	aspen	11/26/97
	VCOOLR	$7,000	$8,400	$6,700	$6,800	$7,600		June, 1996	aspen	11/26/97
Pumps										
	PTOWR	$4,200	$8,000	$2,800	$3,200	$5,500		June, 1996	aspen	11/26/97
	PHOTH2O	$1,000	$3,500	$0	$400	$3,500		price quote from General Pumps		
	PSTEAM	$57,400	$75,100	$57,400	$57,400	$59,300		June, 1996	aspen	11/26/97
	totals	$77,200	$103,100	$66,900	$74,600	$83,500				

DSRP costs

Expendatures

		DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500	cost ref.	
Electrical								
Pumps & Compressors								
kW	RECYCOMP	59	227	5	7	37	ASPEN generated power requirements	
kW	PRESAIR	3282	10414	900	999	4889	ASPEN generated power requirements	
kW	Steam pumps	76	193	30	32	160	ASPEN steam simulations	11/26/97
Light & instruments								
kW	misc.	683	683	683	683	683	20% base case pump & compressor requirements	
TOTAL	kW	4100.4	11517.4	1618.4	1721	5769		
	unit cost $/kWh	0.04	0.04	0.04	0.04	0.04	Self-gen. (Jan. 1990) Peters & Timmeraus	
90 % op	Cost $/yr	$1,293,988	$3,634,615	$510,728	$543,234	$1,820,690		
Cooling Water								
	lbs/hr	149,000	500,000	25,000	62,744	313,720	ASPEN Complete Steam Generation Scheme simulations	
	unit value $/lb	2.6E-05	2.6E-05	2.6E-05	2.6E-05	2.6E-05	Tower (Jan. 1990) Peters & Timmeraus	
90 % op	Cost $/yr	$21,854	$73,336	$3,667	$9,203	$46,014		
Oxygen								
	lbs/hr	0	0	0	0	0		
	unit value $/lb							
	Cost $/yr							
Addtional Employees								
	Engineers	2	2	2	2	2		
	unit cost	$100,000	$100,000	$100,000	$100,000	$100,000		
	Maintenance	2	2	2	2	2		
	unit cost	$70,000	$70,000	$70,000	$70,000	$70,000		
	Cost $/yr	$340,000	$340,000	$340,000	$340,000	$340,000		
Consumed Coal Gas								
	MW lost	7	23	2	3	15	Appendix J	
	unit cost $/MWh	40	40	40	40	40	Self-gen. (Jan. 1990) Peters & Timmeraus	
	Cost $/yr	$2,287,295	$7,259,195	$652,927	$963,138	$4,714,074		
	totals (yearly)	$3,943,137	$11,307,146	$1,507,322	$1,855,574	$6,920,778		

DSRP costs

Benefits

		DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500	Condition	value ref.	date of calc.
Sulfur Recovered									
	lbs/hr	5,840	18,590	1,667	2,460	12,300			
90% op	tons/year	23,037	73,332	6,576	9,704	48,520			11/4/97
	unit value $/ton	50	50	50	50	50	low purity	Chem. Eng. Progress 1996	
	Revenue $/yr	$1,151,852	$3,666,599	$328,791	$485,198	$2,425,991			
Steam Generation									
	lbs/hr	23,200	77,700	6,160	9,800	48,800	950 psia, 441 C		11/4/97
	unit value $/lb	0.0039	0.0039	0.0039	0.0039	0.0039	500 psig, (Jan. 1990)	Peters and Timmeraus	
90% op	Revenue $/yr.	$713,833	$2,390,725	$189,535	$301,533	$1,501,511			
	totals (yearly)	$1,865,685	$6,057,324	$518,326	$786,731	$3,927,501			

	DSRP	DSRP-b	DSRP-c	DSRP-100	DSRP-500
YEARLY COST	$2,077,452	$5,249,823	$988,996	$1,068,843	$2,993,277
EQUIPMENT COSTS	$3,393,129	$7,245,855	$2,039,375	$1,927,612	$6,098,200

AHGP Costs

Equipment

Type	unit	AHG Price	AHG-b Price	AHG-c Price	AHG-100 Price	AHG-500 Price	Mat. of Construction	Purchase date	price ref.	date of calculation
Heat Exchangers										
	HEATX	$64,900	$125,700	$32,900	$39,600	$107,300	SS310 (SS 316)	June, 1996	aspen	12/3/97 AHGPcosts
	RCYHEATR	$102,800	$162,900	$35,300	$60,500	$181,000	SS310 (SS 316)	June, 1996	aspen	1/22/98 steam
	N2-COOLR	$42,000	$72,200	$16,800	$26,500	$66,400	SS304	June, 1996	aspen	1/22/98 steam
Condensers										
	COND	$82,200	$177,000	$41,000	$51,400	$138,500	SS310-heat exchanger	June, 1996	aspen	1/22/98 steam
	LP-COND	$8,200	$11,100	$6,200	$7,000	$10,000	SS310 tank (τ = 1min)	June, 1996	aspen	12/3/97 AHGPcosts
Demister										
	DEMISTR	$53,100	$109,000	$30,600	$35,000	$83,700	SS310 1.5tank (τ =1mir	June, 1996	aspen	12/3/97 AHGPcosts
Compressor										
	CON-COMP	$201,100	$203,300	$200,900	$200,900	$202,100	3 x (Carbon Steel)	June, 1996	aspen	12/3/97 AHGPcosts
	LIFTCOMP	$485,000	$820,000	$161,600	$161,600	$820,000	3 x (Carbon Steel)	June, 1996	aspen mod.	12/3/97 AHGPcosts
	SO2-COMP	$53,900	$66,200	$53,900	$53,900	$1,410,000	3 x (Carbon Steel)	June, 1996	aspen	12/3/97 AHGPcosts
Tanks										
7 days	storage	$125,500	$205,400	$65,000	$80,000	$171,000	SS316	June, 1996	aspen	11/6/97
Reactors										
		$2,939,588	$12,297,497	$1,562,672	$2,177,791	$8,869,074	SS310	June, 1996 (w install)	P&T calc	11/20/97
Pipes										
	pipe lines									
	totals	$4,158,288	$14,250,297	$2,206,872	$2,894,191	$12,059,074				

Equipment -Steam side

Type	unit	AHGP Price	AHGP-b Price	AHGP-c Price	AHGP-100 Price	AHGP-500 Price	Mat. of Construction	date		Purchase price ref.
pumps										
	PTOWR	$3,400	$5,000	$2,800	$2,800	$4,300		June, 1996	aspen	11/26/97 steam
	PSTEAM	$57,400	$63,300	$57,400	$57,400	$59,300		June, 1996	aspen	11/26/97 steam
Heat Exchangers										
	VCOOLR	$7,000	$8,000	$6,700	$6,800	$7,600	shell CS / tube 304	June, 1996	aspen	11/26/97 steam
	totals	$67,800	$76,300	$66,900	$67,000	$71,200				

Expendatures

		AHGP	AHGP-b	AHGP-c	AHGP-100	AHGP-500		cost ref.	
Electrical							1 kW = 1.341 hp		
Pumps & Compressors									
kW	CON-COMP	8	26	2	3	17	ASPEN generated power requirements		1/22/98
kW	LIFTCOMP	13	28	3	5	27	ASPEN generated power requirements		1/22/98
kW	SO2-COMP	38	114	11	16	80	ASPEN generated power requirements		1/22/98
kW	Steam pumps	64	148	28	27	135	ASPEN generated power requirements		
Light & instruments									
kW	misc.	25	25	25	25	25	20% base case pump & compressor requirements		
TOTAL	kW	148	341	68	76	284			
	unit cost $/kWh	0.04	0.04	0.04	0.04	0.04	Self-gen. (Jan. 1990)	Peters & Timmeraus	
90 % op	Cost $/yr	$46,579	$107,485	$21,491	$24,109	$89,490			
Cooling Water									
	lbs/hr	79,200	4,530	434	33,351	166,756			
	unit value $/lb	2.6E-05	2.6E-05	2.6E-05	2.6E-05	2.6E-05	Tower (Jan. 1990)	Peters & Timmeraus	
90 % op	Cost $/yr	$29,041	$1,661	$159	$12,229	$61,146			
Oxygen									
	lbs/hr	4,129	12,536	1,195	1,739	8,694			
	unit value $/ton	$20	$20	$20	$20	$20	Increased O2 plant production	Dr. Roberts	
90 %op	Cost $/yr	$325,753	$989,015	$94,278	$137,175	$685,874			
Additional Employees									
	Engineers	3	3	3	3	3			
	unit cost	$100,000	$100,000	$100,000	$100,000	$100,000			
	Maintenance	2	2	2	2	2			
	unit cost	$70,000	$70,000	$70,000	$70,000	$70,000			
	Cost $/yr	$440,000	$440,000	$440,000	$440,000	$440,000			
Consumed Coal Gas									
	MW lost	1.506	4.580	0.436	0.634	3.172	Appendix J		
	unit cost $/MWh	40	40	40	40	40	Self-gen. (Jan. 1990)	Peters & Timmeraus	
	Cost $/yr	$475,257	$1,445,338	$137,591	$200,075	$1,001,007			
	totals (yearly)	$1,316,631	$2,983,500	$693,519	$813,588	$2,277,517			

Benefits

		AHGP	AHGP-b	AHGP-c	AHGP-100	AHGP-500	Condition	value ref.	date of calc.
Sulfur Recovered									
	lbs/hr	5,731	17,440	1,593	2,413	12,067			
90% op	tons/year	22,607	68,796	6,284	9,520	47,599			11/4/97
	unit value $/ton	50	50	50	50	50	low purity	Chem. Eng. Progress 1996	
	Revenue $/yr	$1,130,354	$3,439,778	$314,195	$475,992	$2,379,960			
Steam Generation									
	lbs/hr	19,400	59,000	5,650	8,169	40,847	950 psia, 441 C		11/4/97
	unit value $/lb	0.0039	0.0039	0.0039	0.0039	0.0039	500 psig, (Jan. 1990)	Peters and Timmeraus	
90% op	Revenue $/yr.	$596,912	$1,815,351	$173,843	$251,360	$1,256,798			
	totals (yearly)	$1,727,266	$5,255,129	$488,038	$727,352	$3,636,758			

	AHGP	AHGP-b	AHGP-c	AHGP-100	AHGP-500
YEARLY COST	-$410,635	-$2,271,630	$205,481	$86,236	-$1,359,241
EQUIPMENT COSTS	$4,226,088	$14,326,597	$2,273,772	$2,961,191	$12,130,274

AHGP Costs

Equipment

Type	unit	AHG Price	AHG-b Price	AHG-c Price	AHG-100 Price	AHG-500 Price	Mat. of Construction	Purchase date	price ref.	date of calculation
Heat Exchangers										
	HEATX	$64,900	$125,700	$32,900	$39,600	$107,300	SS310 (SS 316)	June, 1996	aspen	12/3/97 AHGPcosts
	RCYHEATR	$102,800	$162,900	$35,300	$60,500	$181,000	SS310 (SS 316)	June, 1996	aspen	1/22/98 steam
	N2-COOLR	$42,000	$72,200	$16,800	$26,500	$66,400	SS304	June, 1996	aspen	1/22/98 steam
Condensers										
	COND	$82,200	$177,000	$41,000	$51,400	$138,500	SS310-heat exchanger	June, 1996	aspen	1/22/98 steam
	LP-COND	$8,200	$11,100	$6,200	$7,000	$10,000	SS310 tank (τ = 1min)	June, 1996	aspen	12/3/97 AHGPcosts
Demister										
	DEMISTR	$53,100	$109,000	$30,600	$35,000	$83,700	SS310 1.5tank (τ =1min	June, 1996	aspen	12/3/97 AHGPcosts
Compressor										
	CON-COMP	$201,100	$203,300	$200,900	$200,900	$202,100	3 x (Carbon Steel)	June, 1996	aspen	12/3/97 AHGPcosts
	LIFTCOMP	$485,000	$820,000	$161,600	$161,600	$820,000	3 x (Carbon Steel)	June, 1996	aspen mod.	12/3/97 AHGPcosts
	SO2-COMP	$53,900	$66,200	$53,900	$53,900	$1,410,000	3 x (Carbon Steel)	June, 1996	aspen	12/3/97 AHGPcosts
Tanks										
7 days	storage	$125,500	$205,400	$65,000	$80,000	$171,000	SS316	June, 1996	aspen	11/6/97
Reactors										
		$2,939,588	$12,297,497	$1,562,672	$2,177,791	$8,869,074	SS310	June, 1996 (w install)	P&T calc	11/20/97
Pipes										
	pipe lines									
	totals	$4,158,288	$14,250,297	$2,206,872	$2,894,191	$12,059,074				

AHGP Costs

Equipment -Steam side

Type	unit	AHGP Price	AHGP-b Price	AHGP-c Price	AHGP-100 Price	AHGP-500 Price	Mat. of Construction	date		Purchase price ref.
pumps										
	PTOWR	$3,400	$5,000	$2,800	$2,800	$4,300		June, 1996	aspen	11/26/97 steam
	PSTEAM	$57,400	$63,300	$57,400	$57,400	$59,300		June, 1996	aspen	11/26/97 steam
Heat Exchangers										
	VCOOLR	$7,000	$8,000	$6,700	$6,800	$7,600	shell CS / tube 304	June, 1996	aspen	11/26/97 steam
	totals	$67,800	$76,300	$66,900	$67,000	$71,200				

AHGP Costs

Expendatures

		AHGP	AHGP-b	AHGP-c	AHGP-100	AHGP-500	cost ref.	
Electrical							1 kW = 1.341 hp	
Pumps & Compressors								
kW	CON-COMP	8	26	2	3	17	ASPEN generated power requirements	1/22/98
kW	LIFTCOMP	13	28	3	5	27	ASPEN generated power requirements	1/22/98
kW	SO2-COMP	38	114	11	16	80	ASPEN generated power requirements	1/22/98
kW	Steam pumps	64	148	28	27	135	ASPEN generated power requirements	
Light & instruments								
kW	misc.	25	25	25	25	25	20% base case pump & compressor requirements	
TOTAL	kW	148	341	68	76	284		
	unit cost $/kWh	0.04	0.04	0.04	0.04	0.04	Self-gen. (Jan. 1990) Peters & Timmeraus	
90 % op	Cost $/yr	$46,579	$107,485	$21,491	$24,109	$89,490		
Cooling Water								
	lbs/hr	79,200	4,530	434	33,351	166,756		
	unit value $/lb	2.6E-05	2.6E-05	2.6E-05	2.6E-05	2.6E-05	Tower (Jan. 1990) Peters & Timmeraus	
90 % op	Cost $/yr	$29,041	$1,661	$159	$12,229	$61,146		
Oxygen								
	lbs/hr	4,129	12,536	1,195	1,739	8,694		
	unit value $/ton	$20	$20	$20	$20	$20	Increased O2 plant production Dr. Roberts	
90 %op	Cost $/yr	$325,753	$989,015	$94,278	$137,175	$685,874		
Additional Employees								
	Engineers	3	3	3	3	3		
	unit cost	$100,000	$100,000	$100,000	$100,000	$100,000		
	Maintenance	2	2	2	2	2		
	unit cost	$70,000	$70,000	$70,000	$70,000	$70,000		
	Cost $/yr	$440,000	$440,000	$440,000	$440,000	$440,000		
Consumed Coal Gas								
	MW lost	1.506	4.580	0.436	0.634	3.172	Appendix J	
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AHGP Costs

Benefits

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	AHGP	AHGP-b	AHGP-c	AHGP-100	AHGP-500
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EQUIPMENT COSTS	$4,226,088	$14,326,597	$2,273,772	$2,961,191	$12,130,274

Appendix O
Reaction Data Obtained from RTI

The following data was obtained during correspondence with RTI.

DSRP reactions at 300 psi

Reaction	ΔH at 550°C (J/mol)	ΔH at 650°C (J/mol)	ΔH at 750°C (J/mol)
0.5 SO_2 + H_2 = (1/4)S_2 + H_2O	-65128	-65795	-66436
0.5 SO_2 + CO = (1/4)S_2 + CO_2	-101938	-101629	-101295

ZnO + H_2S(g) = ZnS + H_2O(g)

Temp. °C	ΔH kcal	ΔS cal	ΔG kcal	K
400	-17.079	-0.071	-17.031	3.387E+5
500	-17.056	-0.040	-17.025	6.502E+4
600	-17.047	-0.029	-17.022	1.824E+4
700	-17.050	-0.032	-17.019	6.645E+3

ZnS + 1.5 O_2(g) = ZnO + SO_2(g)

Temp. °C	ΔH kcal	ΔS cal	ΔG kcal	K
500	-107.110	-18.940	-92.467	1.381E+26
550	-107.135	-18.971	-91.519	1.999E+24
600	-107.155	-18.995	-90.570	4.694E+22
650	-107.172	-19.013	-89.620	1.654E+21
700	-107.185	-19.027	-88.669	8.220E+19
750	-107.195	-19.038	-87.717	5.474E+18
800	-107.204	-19.046	-86.765	4.692E+17