

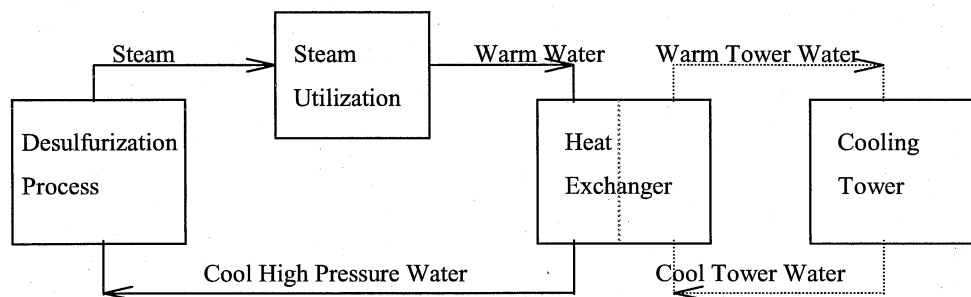
VI. ADDITIONAL PROCESS CONSIDERATIONS

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 ($\sim 90^{\circ}\text{C}$), 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°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°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

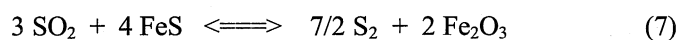
Type 310 stainless steel (25%Cr - 20%Ni) should be used for the construction of equipment that contacts sulfur species. Type 310 stainless steel (SS 310) will be more durable than type 316 stainless steel (SS 316) (17%Cr - 8% Ni - 2%Mo). Higher chromium content gives SS310 greater oxidation resistance, 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 SO₂ production and consumption in the AHGP appears to be particularly difficult. The difficulty arises from the reaction of FeS with SO₂ to form elemental sulfur. The reaction's equilibrium varies significantly with temperature. If the reactants are too thermodynamically favored, less SO₂ will be consumed than expected. However, SO₂ production will remain constant (sorbent oxidation being driven to completion by oxygen). Thus, if the reaction:



does not reach design conversions, SO₂ flow will increase and sulfurized sorbent will be returned to the desulfurization reactor. With SO₂ already present in great excess the increased SO₂ 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 SO₂. Replenishing depleted SO₂ levels can easily be accomplished by increasing the oxygen feed. Excess oxygen will convert elemental sulfur into SO₂.

Preventing the build up of impurities in the SO₂ loop contributes to the complexity of the AHGP. Venting a portion of the loop is undesirable since it contains mostly SO₂. Venting would release SO₂, 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 SO₂.

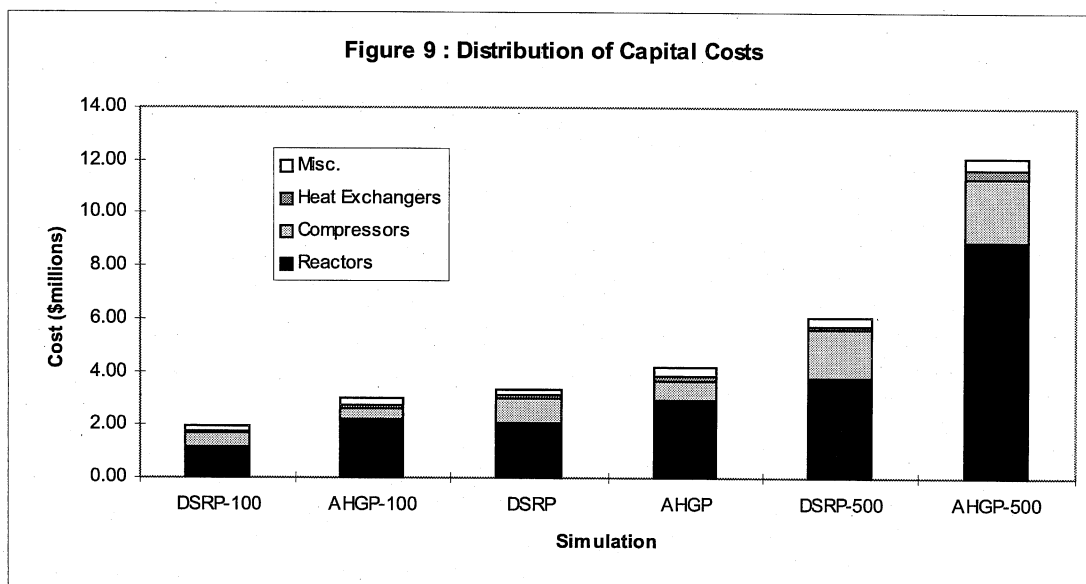
VII. ECONOMIC ANALYSIS

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.

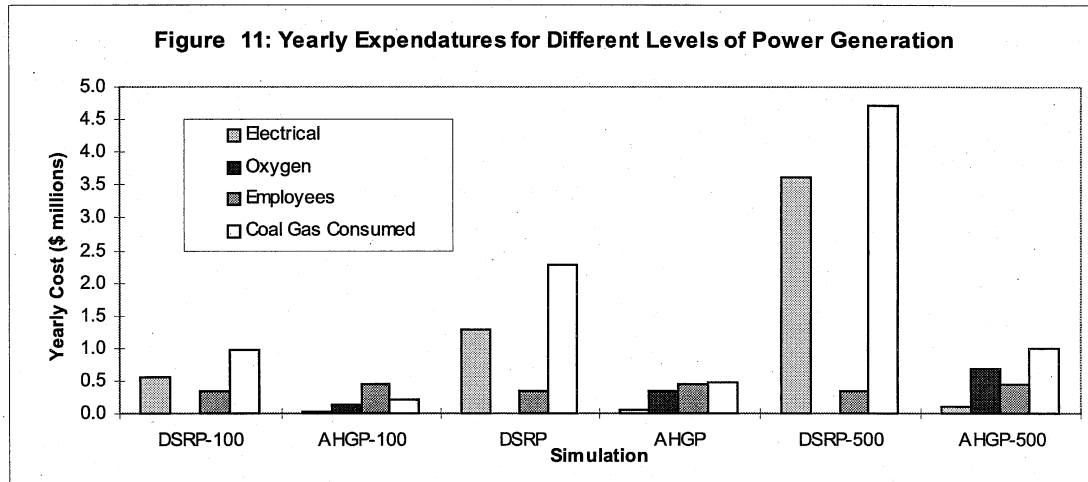
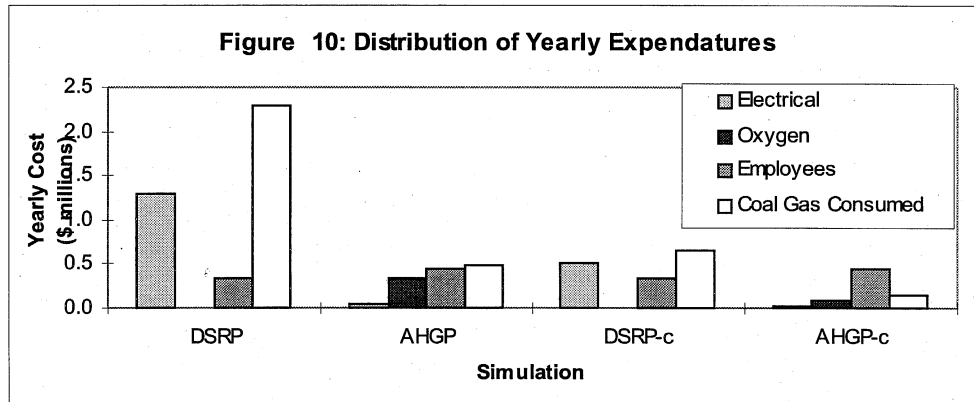
Another piece of equipment not costed by ASPEN is the PRESAIR - air compressor used in the DSRP. PRESAIR costs were determined by scaling a price quote for the Ingersoll-Rand Centac air compressor. The Centac Model 2CV23M3EPPF, capable of raising 2,250 acfm to 280 psia, was quoted at \$241,000. Extrapolation 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.



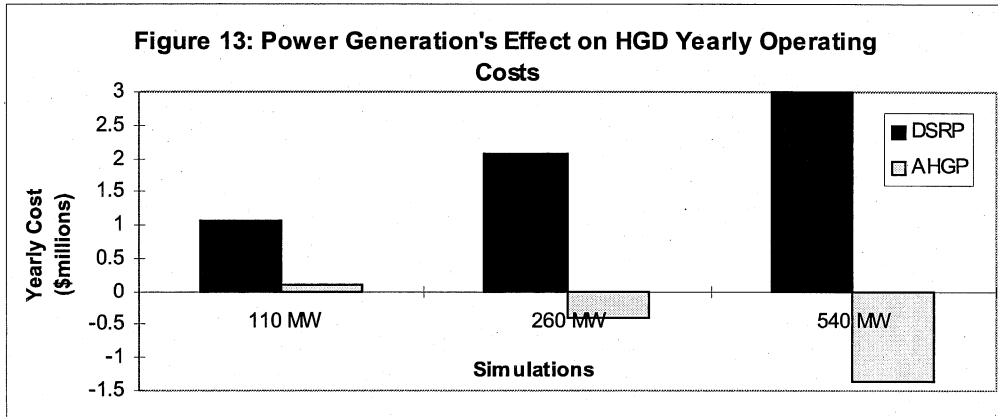
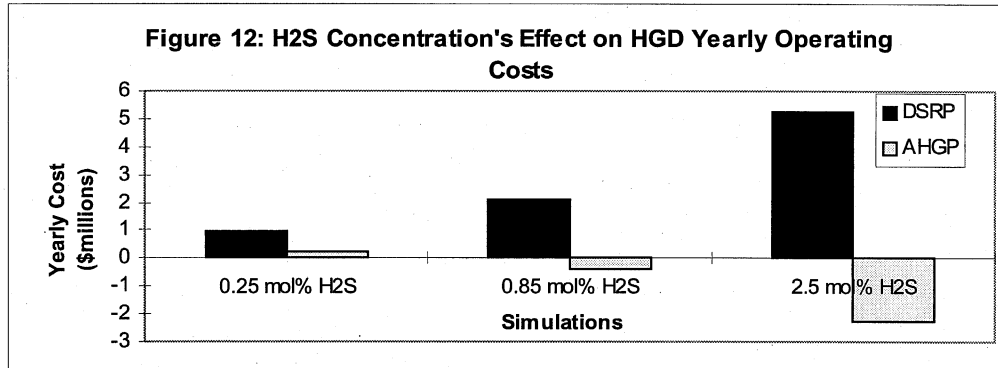
There are additional capital costs not included in this report, two of which, piping costs and sorbent/catalyst costs, will probably be significant. There will be other expenses, like additional office space for employees, which are site dependent. The site dependent expenses should not have a significant 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

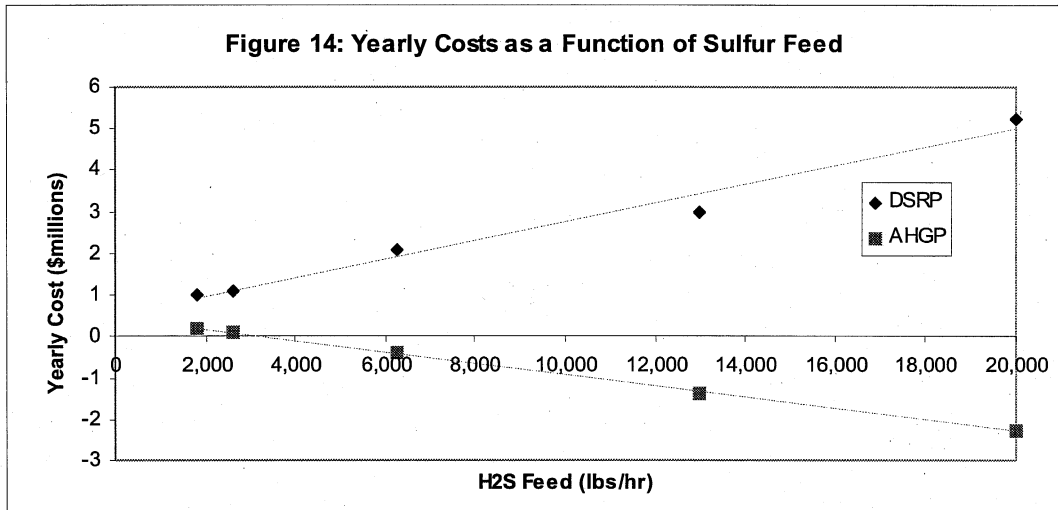
The AHGP has a lower yearly operating cost than the DSRP. Figures 10 and 11 show the distribution of the major yearly expenditures for both processes.



The bases cases (DSRP and AHGP) have coal gas feeds containing 0.85 mol% H₂S 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% H₂S). 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 $(\text{[DSRP cost - AHGP cost]} / \text{AHGP cost})$ decrease. This decrease indicates that the competitive advantage of the AHGP is smaller for cleaning a coal gas stream containing a low H₂S 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.



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 were 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.



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. There 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°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°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°C, before its reuse. The cost of the tower water is $\$2.6 \times 10^{-5}/\text{lb}$ (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 SO₂ production and consumption must be balanced. Furthermore, the purity of the SO₂ 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 (H₂ 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 H₂ lost during HGD, and calculating the value of the energy that the CO and H₂ could have produced. Calculation of power generation is described in Appendix J.

The DSRP consumes substantially more coal gas than 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.

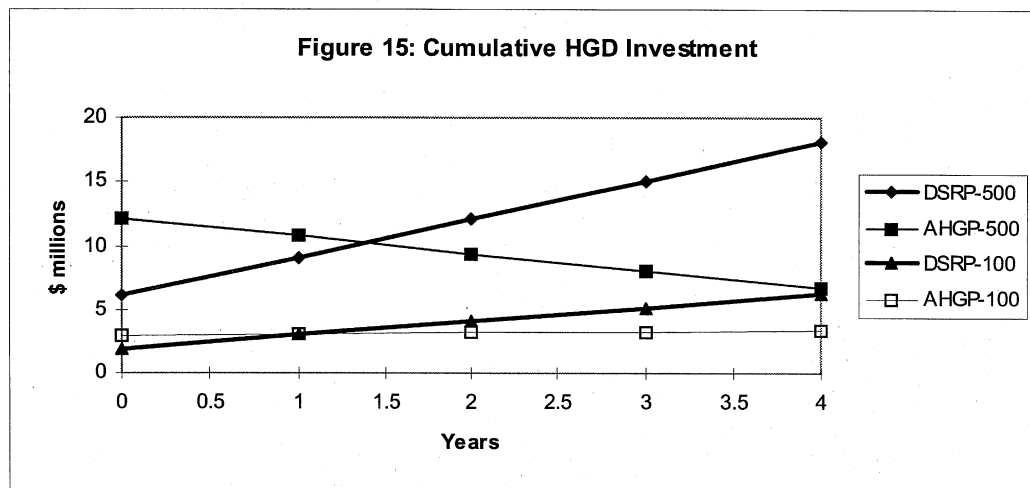


Figure 15 shows that despite an higher initial investment, within two years the AHGP can financially outperform the DSRP.

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 than 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.

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Appendix A
Calculation of the SO₂ Circulation Rate for AHGP

SO₂ 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. Al₂O₃ 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 SO₂ circulation rate necessary to provide a 2.5 cm/s upwards velocity is then calculated. Calculation results follow:

SO2 Regenerator Sizing - Commercial Embodiment

	AGHP (SO2 Regen) Case E-2	AHGP-b (SO2 Regen) Case E-2	AHGP-c
Givens:			
Sorbent circulation rate, lb/hr	166010	496000	48000
Sorbent bulk density, lb/ft ³	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 ft ³ /hr	79,813	238,462	23,077
Calculated values:			
Hold-up volume, ft ³	2660	7949	769
Diameter, ft	18	32	9.9
X-section area, ft ²	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/ft ³	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 °F
Water	Water	250 - 500
Gases	Water	2 - 50
Heaters		
Hot Fluid	Cold Fluid	Overall U_D , BTU/hr ft ² °F
Steam	Water	200 - 700
Steam	Gases	5 - 50

Values above found in Kern (1950).

Fluid combination	U, BTU/hr ft ² °F
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:

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

The catalyst size is 160 micron.

$$D_p = 1.6 \times 10^{-4} \text{ m} \quad \rho_p = 1.2 \text{ g/cm}^3 \quad g = 9.8 \text{ m/s}^2$$

Bulk samples of the catalyst have a density (ρ_{bulk}) of 0.9 g/cm³. 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).

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

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

$$\rho = 0.50 \text{ lb/ft}^3 \times (1,000 \text{ gr}) / (2.205 \text{ lb}) \times (1 \text{ ft}^3) / (30.48 \text{ cm})^3 = 0.008 \text{ g/cm}^3 = 8 \text{ kg/m}^3$$

Inserting the values gives:

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

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 be assumed to have the properties of nitrogen (N_2 represents over 50 wt% of reactor gas).

Reactor conditions $T = 600^\circ\text{C}$ 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.

Equations

$$(\mu - \mu^0)/(\mu^0 \rho) = A P_r^{1.5} / [B P_r + (1 + C P_r^D)^{-1}]$$

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

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

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

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

Nitrogen Properties

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

$$\mu^0 = \mu (1 \text{ atm}, 873 \text{ K}) = 3.8 \times 10^8 \text{ Poise}$$

$$\text{And for nonpolar molecules: } \rho = 1$$

Calculated Values

$$A = 0.001615 \quad B = 0.0164 \quad C = 0.01909 \quad D = 0.4269$$

$$(\mu - \mu^0) / (\mu^0) = 6.7498 \times 10^{-4} / 0.9945 = 6.787 \times 10^{-4}$$

$$\mu = 3.8 \times 10^8 + (3.8 \times 10^8)(6.787 \times 10^{-4}) \text{ Poise}$$

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

The Reynolds number is can now be expressed:

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

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

And our velocity equation is:
$$v_{ss} = \sqrt{\frac{0.3136 \frac{\text{m}^2}{\text{s}^2}}{C_D}}$$

First Iteration, take $v_{ss} = 0.9 \text{ m/s}$ then $N_{Re} = 30$ (above equation)

For the above Reynolds number $C_D = 2.4$ (Fig. 6.3-1 in Bird)

Velocity equation gives $v_{ss} = 0.36 \text{ m/s}$

The calculations are repeated.

Second iteration:	$v_{ss} = 0.36 \text{ m/s}$	$N_{Re} = 12$
	$C_D = 4.2$	$v_{ss} = 0.27 \text{ m/s}$

Third iteration:	$v_{ss} = 0.27 \text{ m/s}$	$N_{Re} = 9.09$
	$C_D = 4.9$	$v_{ss} = 0.252 \text{ m/s}$

Fourth iteration:	$v_{ss} = 0.252 \text{ m/s}$	$N_{Re} = 8.49$
	$C_D = 5.13$	$v_{ss} = 0.247 \text{ m/s}$

Fifth iteration:	$v_{ss} = 0.247 \text{ m/s}$	$N_{Re} = 8.33$
	$C_D = 5.19$	$v_{ss} = 0.246 \text{ m/s}$

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 which 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°C and the catalyst is cooled to 500°C.

Catalyst Properties

The DSRP reactor catalyst is a porous aluminum oxide catalyst modeled as Al₂O₃. Catalyst density at ambient conditions is 56.18 lb/ft³. 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 Al₂O₃ and 0.00095 lb of N₂ have a combined density of 55.6 lb/ft³.

The similar densities allow us to assume that there is roughly 0.00095 lb of nitrogen present for every 1 lb of solid Al₂O₃ (at ambient conditions). That quantity of nitrogen occupies 0.0132 ft³ (at ambient conditions). This represents the catalyst void volume and is expected to remain constant.

$$V_{\text{void}} = 0.0132 \text{ ft}^3 / 1 \text{ lb Al}_2\text{O}_3$$

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

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

Therefore the mass of gas (in the settled catalyst) per pound Al₂O₃ can be calculated.

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

The heat transfer requirements for cooling Al₂O₃ were then simulated (including cooling nitrogen contained in the catalyst voids).

$$Q(600^\circ\text{C} \rightarrow 500^\circ\text{C}) = -51.239 \text{ BTU/lb Al}_2\text{O}_3$$

Calculation of necessary catalyst circulation rate:

$$(\text{circulation rate } \{\text{lb/hr}\}) = (Q_{\text{DSRP}}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3)$$

$$\begin{aligned} \text{DSRP} \quad (\text{circulation rate } \{\text{lb/hr}\}) &= (-15,340,000 \text{ BTU/hr}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3) \\ &= \mathbf{300,000 \text{ lb Al}_2\text{O}_3 / \text{hr}} \end{aligned}$$

$$\begin{aligned} \text{DSRP-b} \quad (\text{circulation rate } \{\text{lb/hr}\}) &= (-51,320,000 \text{ BTU/hr}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3) \\ &= \mathbf{1,000,000 \text{ lb Al}_2\text{O}_3 / \text{hr}} \end{aligned}$$

$$\begin{aligned} \text{DSRP-c} \quad (\text{circulation rate } \{\text{lb/hr}\}) &= (-4,029,000 \text{ BTU/hr}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3) \\ &= \mathbf{79,000 \text{ lb Al}_2\text{O}_3 / \text{hr}} \end{aligned}$$

$$\begin{aligned} \text{DSRP-100} \quad (\text{circulation rate } \{\text{lb/hr}\}) &= (-6,459,000 \text{ BTU/hr}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3) \\ &= \mathbf{130,000 \text{ lb Al}_2\text{O}_3 / \text{hr}} \end{aligned}$$

$$\begin{aligned} \text{DSRP-500} \quad (\text{circulation rate } \{\text{lb/hr}\}) &= (-31,370,000 \text{ BTU/hr}) / (-51.239 \text{ BTU/lb Al}_2\text{O}_3) \\ &= \mathbf{610,000 \text{ lb Al}_2\text{O}_3 / \text{hr}} \end{aligned}$$

Appendix E
Process Flowsheets and Stream Summaries

Direct Sulfur Recovery Process Simulations

DSRP (base case)	0.85 mole% H ₂ S	260 MW generated
DSRP-b	2.50 mole% H ₂ S	260 MW generated
DSRP-c	0.25 mole% H ₂ S	260 MW generated
DSRP-100*	0.85 mole% H ₂ S	110 MW generated
DSRP-500	0.85 mole% H ₂ S	540 MW generated

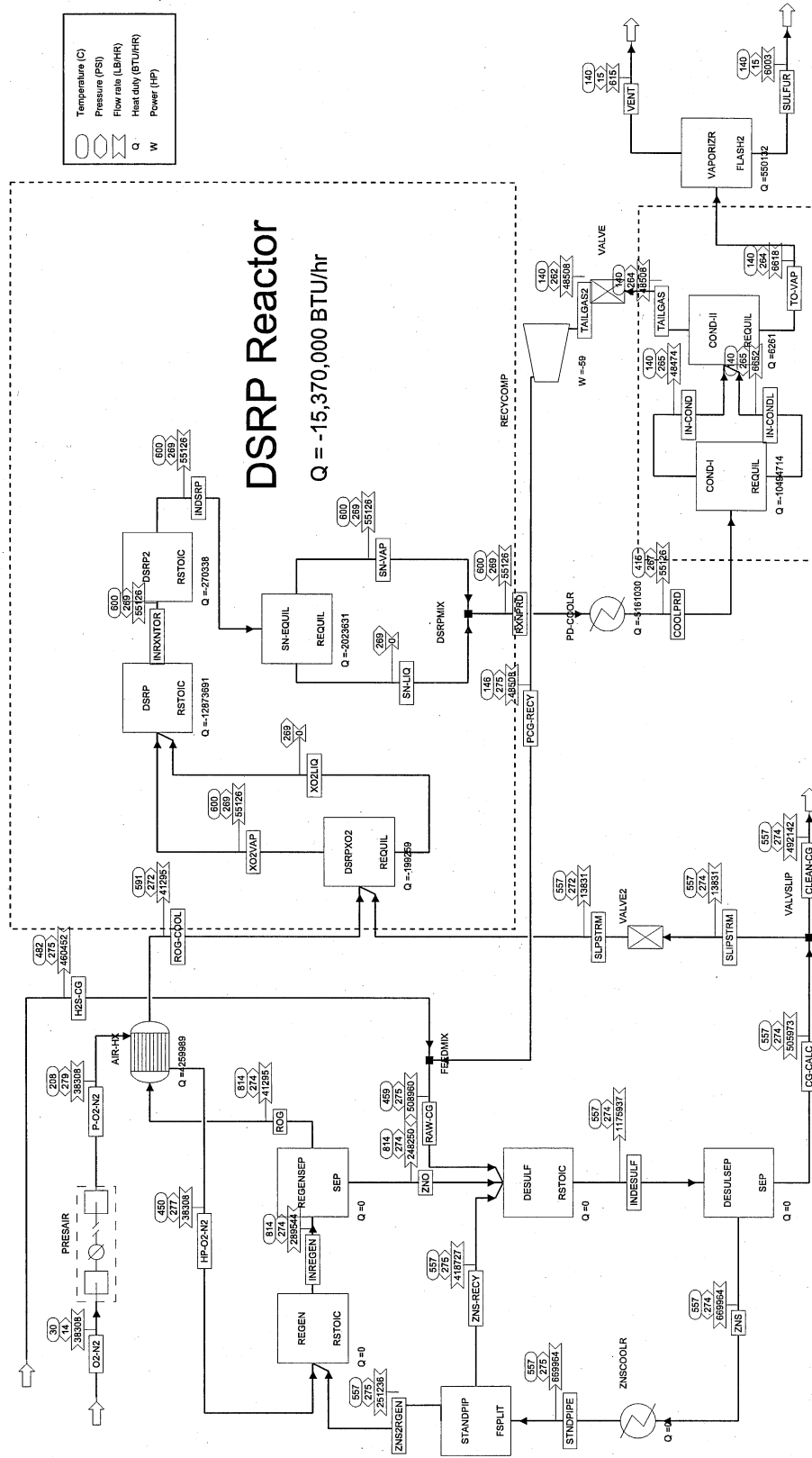
Advanced Hot Gas Process Simulations

AHGP (base case)	0.85 mole% H ₂ S	260 MW generated
AHGP-b	2.50 mole% H ₂ S	260 MW generated
AHGP-c	0.25 mole% H ₂ S	260 MW generated
AHGP-100*	0.85 mole% H ₂ S	110 MW generated
AHGP-500*	0.85 mole% H ₂ S	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



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Display ALLS Units: Format: SOLI DS To Phas	TREAMS From DS To Phas	CG-CALC DESULSEP VALVS VAPOR	CLEAN-CG VALVS VAPOR	COOLPRD PD-COOLR COND-I VAPOR	H2S-CG FEEDMIX VAPOR	HP-O2-N2 AIR-HX REGEN VAPOR	IN-COND COND-I COND-H VAPOR	IN-COND COND-I COND-H LIQUID	INDESULF DESULSEP VAPOR	INDSRP DSRP2 SN-EQUIL VAPOR	INREGEN REGEN VAPOR	INRXNTOR DSRP VAPOR	IO2-N2 PRESAIR VAPOR	Display ALLS Units: Format: SOLI DS To Phas	TREAMS From DS To Phas
Temperature [C]	557.2	557.2	557.2	450	482.2	450	140	140	557.2	600	814.4	600	30	Temperature [C]	
Pressure [PSI]	274.4	274.4	274.4	276.9	275	276.9	264.6	264.6	274.4	266.6	273.6	266.6	13.7	Pressure [PSI]	
Mass VFrac	1	1	1	0	0	0	1	0	0.43	0	0.143	0	1	Mass VFrac	
*** ALL PHAS ES ***	0	0	0	0	0	0	0	0	0.57	0	0.857	0	0	*** ALL PHAS ES ***	
Mass Flow [LB/HR]	505972.781	492141.531	492141.531	55125.902	460451.688	38308.016	6651.654	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	1.33E+06	87129.523	1.15E+06	67432.328	49166.316	481953	1.37E+06	112094.914	96496.891	111308.695	567372.75	Volume Flow [CUFT/HR]	
Enthalpy [BTU/HR]	-1.15E+09	-1.11E+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	0.37	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]	63.71	61.968	61.968	63.71	63.71	63.71	0.18	0.18	63.71	63.879	11909.276	1787.733		Mass Flow [LB/HR]	
O2S	6.82	6.634	6.634	72.073	6270.481	71.958	0.116	0.116	6.82	72.072	72.072	72.072		O2S	
H2O	78082.43	75947.969	75947.969	4963.325	70525.25	4175.116	788.209	788.209	78082.43	4963.325	4963.325	4963.325		H2O	
S2	< 0.001	< 0.001	< 0.001	< 0.001	< 0.001	< 0.001	0.212	< 0.001	0.212	5861.615	3273.283	3273.283		S2	
S6	0.39	0.38	0.38	1891.255	1891.255	1891.255	0.965	1890.29	0.39	0.011	0.011	0.011		S6	
S8	2.764	2.688	2.688	1339.271	1339.271	1339.271	2.025	3968.209	2.764	0.076	0.076	0.076		S8	
CO	218164.266	212200.516	212200.516	2.262	218162	218162	2.262	< 0.001	218164.266	2.262	2.262	2.262		CO	
CO2	130332.672	126769.898	126769.898	12929.425	117407.195	12929.425	4.181	130332.672	12929.425	12929.425	10560.996	10560.996		CO2	
H2	11766.221	11444.579	11444.579	0.85	11766.37	11766.37	0.85	< 0.001	11766.221	0.85	0.85	0.85		H2	
O2	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588	8922.588		O2	
N2	67553.203	65706.57	65706.57	31232.063	36321.383	29385.428	0.256	67553.203	31232.063	31232.063	29385.428	31232.063		N2	
COS	0.323	0.314	0.314	0.323	0.323	0.323	0.323	< 0.001	0.323	0.323	0.323	0.323		COS	
ZNO											15129.819	3233.392		ZNO	
ZNS											36055			ZNS	
AL2O3											197064.938			AL2O3	

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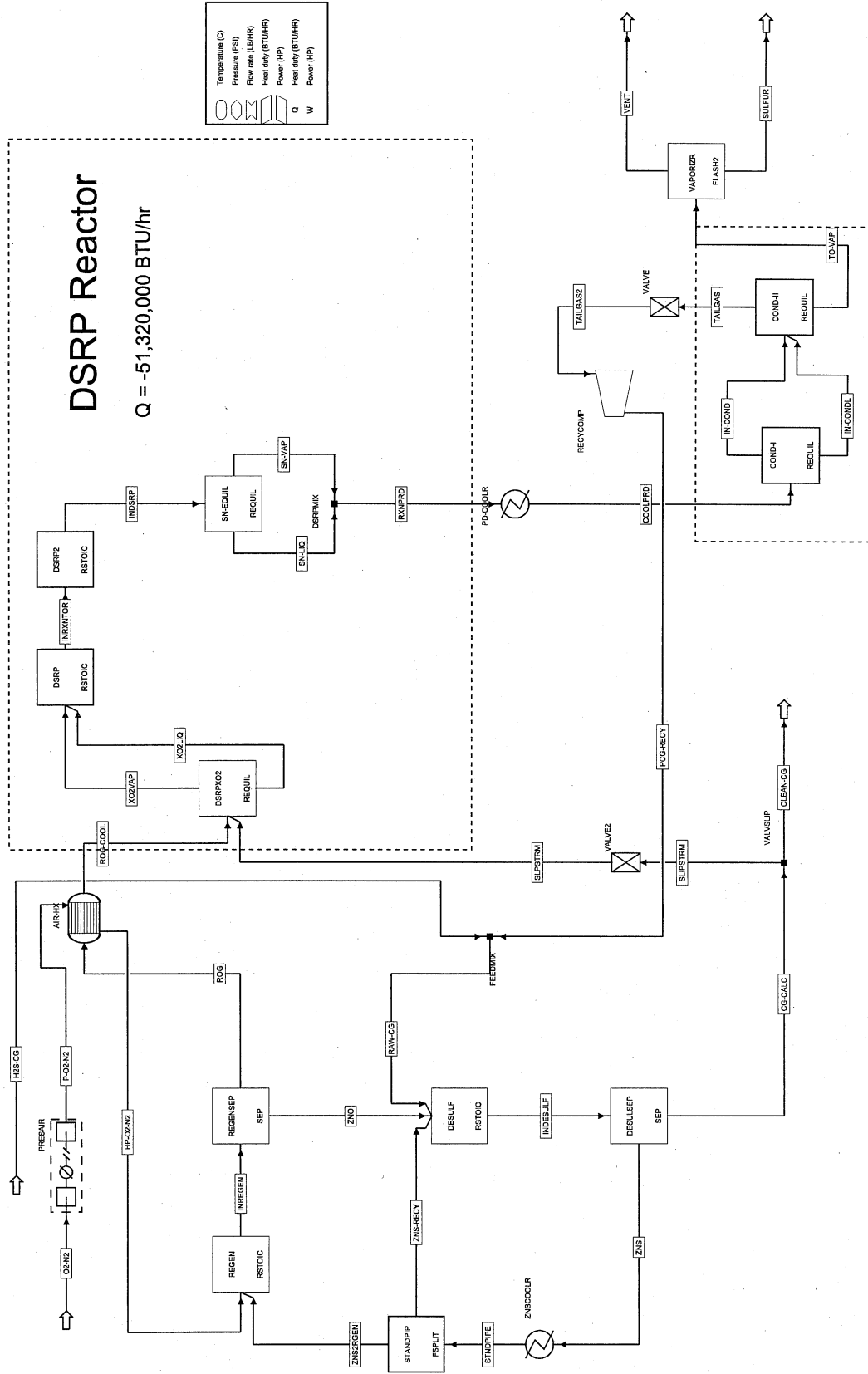
Display ALLS	TREAMS From DS To Phas	P-O2-N2 PRESAIR AIR-HX VAPOR	PCG-RECY RECYCOMP FEEDMIX VAPOR	RAW-CG FEEDMIX DESULF VAPOR	ROG REGENSEP AIR-HX VAPOR	ROG-COOL AIR-HX DSRPX02 VAPOR	RXNPRD DSRPMIX PD-COOLR VAPOR	SLIPSTRM VALVSLIP VALVE2 VAPOR	SLPSTRM VALVE2 DSRPX02 VAPOR	SN-LIQ SN-EQUIL DSRPMIX MISSING	SN-VAP SN-EQUIL DSRPMIX VAPOR	STNDPIPE ZNSCOOLR STANDPIP MISSING	SULFUR VAPORIZR LIQUID	Display ALLS Units: Format: SOLI	TREAMS From DS To Phas
Pressure [PSI]		207.6	146.5	458.8	814.4	590.7	600	557.2	557.2		288.6	600	557.2	14.0 Temperature [C]	
Mass VFrac		278.9	275	275	273.6	271.6	268.6	274.4	271.6		1	1	275	14.7 Pressure [PSI]	
Mass SFrac		0	0	0	0	0	0	0	0		0	0	0	0 Mass VFrac	
*** 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	668963.875	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]		3.08E+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]			63.71	63.71	11909.276	11909.276	63.879	1.742	1.742		63.879		0.001	Mass Flow [LB/HR]	
O2S			71.964	6342.445			72.073	0.186	0.186		72.072		<0.001	O2S	
H2S			4208.244	74733.492			4963.325	2134.462	2134.462		4963.325		148.193	H2S	
H2O			<0.001	<0.001			2631.176	trace	trace		2631.176		0.212	H2O	
S2			0.39	0.39			1891.255	0.011	0.011		1891.255		724.698	S2	
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			12825.476	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		8922.568											trace	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														ZNS	
AL2O3														AL2O3	
														144457.359	
														525506.5	

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Display ALLS	TREAMS	TAILGAS	TAILGAS2	TO-VAP	VENT	XO2LIQ	XO2VAP	ZNO	ZNS	ZNS-RECY	ZNSRGEN	TREAMS
Units:	From DS To Phas	COND-II VALVE RECYCOMP VAPOR	COND-II VALVE RECYCOMP VAPOR	COND-II VAPORIZER LIQUID	VAPORIZER VAPOR	DSRP MISSING	DSRP VAPOR	DESULF MISSING	DESULF MISSING	STANDPIP MISSING	STANDPIP MISSING	From DS To Phas
Format: SOLI	DS To Phas	VAPOR	VAPOR	VAPOR	VAPOR	MISSING	VAPOR	MISSING	MISSING	MISSING	MISSING	Format: SOLI
Temperature [C]	[C]	140	139.9	140	140	140	600	814.4	557.2	557.2	557.2	Temperature [C]
Pressure [PSI]	[PSI]	264.4	262	264.4	14.7	288.6	273.6	274.4	274.4	275	275	Pressure [PSI]
Mass VFrac		1	0	0	0	0	1	0	0	0	0	Mass VFrac
Mass SFrac		0	0	0	0	0	0	1	1	1	1	Mass SFrac
*** ALL PHAS	ES ***											*** ALL PHAS
Mass Flow [LB/HR]	[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]	[CUFT/HR]	49255.441	49703.582	46.323	18288.283	0	118222.719	1195.906	3189.083	1993.177	1195.906	Volume Flow [CUFT/HR]
Enthalpy [BTU/HR]	[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]	[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]	[LB/HR]	63.71	63.71	0.17	0.169		11911.018					Mass Flow [LB/HR]
O2S		71.964	71.964	0.109	0.109		0.186					O2S
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	54171.508	ZNS
AL2O3							197064.938	525506.5	328441.963	197064.938	197064.938	AL2O3

DSRP-b - based Desulfurization

12/20/97 DSRP-b



High Pressure Condenser Q = -35,100,000 BTU/hr

DSRP-b 12/20/97 pg1

Display ALLS Units: Format: SOLI Temperature [C] Pressure [PSI]	TREAMS From DS To Phas	CG-CALC DESULSEP VALVSLIP VAPOR	CLEAN-CG VALVSLIP VAPOR	COOLPRD PD-COOLR COND-I VAPOR	H2S-CG FEEDMIX VAPOR	HP-O2-N2 AIR-HX REGEN VAPOR	IN-COOND COND-I COND-II VAPOR	IN-CONDL COND-I COND-II LIQUID	INDESULF DESULSEP VAPOR	INDSRP DSRP2 SN-EQUIL VAPOR	INREGEN REGEN REGENSEP VAPOR	INRXANTOR DSRP DSRP2 VAPOR	O2-N2 PRESAIR VAPOR	Display ALLS Units: Format: SOLI Temperature [C] Pressure [PSI]	TREAMS From DS To Phas
647.5 273.6 1 0	647.5 273.6 1 0	647.5 273.6 1 0	482.2 275 1 0	415 265.6 1 0	450 277.4 1 0	450 277.4 1 0	140 263.6 0 0	140 263.6 0 0	647.5 273.6 287.6 600	810 272.6 0.099 0.901	600 267.6 1 0	600 267.6 1 0	30 13.7 Pressure Mass SFrac	30 13.7 Temperature [C] Mass SFrac	
655189.75 184610.219	655189.75 184610.219	601554.75 1733330	501234.031 1244680	184610.219 294846.875	121500 213483.484	163323.219 166528.75	21287.002 159.053	3324180 1900580	184607.219 378670.75	1322550 307882.375	184607.719 378670.719	121500 1799510	ES *** Mass Flow [LB/HR] Volume Flow [CUFT/HR] Enthalpy [BTU/HR] Density [LB/CUFT] Mass Flow [LB/HR]	ES *** Mass Flow [LB/HR] Volume Flow [CUFT/HR] Enthalpy [BTU/HR] Density [LB/CUFT] Mass Flow [LB/HR]	
1887870 -1.38E+09 0.347	1887870 -1.24E+09 0.347	1733330 -1.24E+09 0.347	1244680 -1.16E+09 0.403	294846.875 -2.24E+09 0.628	213483.484 2.32E+07 0.589	166528.75 -2.43E+08 0.981	159.053 -1.61E+07 133.636	1900580 -1.60E+10 1.749	378670.75 -2.00E+08 0.468	-6.55E+09 4.296	-1.99E+08 0.491	1799510 2.51E+05	[CUFT/HR] Enthalpy [BTU/HR] Density [LB/CUFT] Mass Flow [LB/HR]	[CUFT/HR] Enthalpy [BTU/HR] Density [LB/CUFT] Mass Flow [LB/HR]	
194.606 62.453 99935.453	194.606 62.453 99935.453	178.675 57.341 91754.563	178.675 57.341 91754.563	195.105 240.342 17155.324	194.579 239.971 14470.959	194.579 239.971 14470.959	0.528 0.371 2684.464	194.606 62.453 99935.453	195.105 240.342 17155.324	37772.176 17155.324	5444.741 240.342 17155.324	195.105 240.342 17155.324	O2S H2O H2O S2	O2S H2O H2O S2	
1.28 8.967 231203.391	1.28 8.967 231203.391	1.175 212276.672	5885.709 4053.262	5885.709 4053.262	1.28 8.967	3.086 6.889	5882.633 12704.021	1.28 8.967	0.105 0.734	18593.258 17155.324	10711.036 0.105	0.105 0.734	S6 S8 CO	S6 S8 CO	
167878.578 12471.098	167878.578 12471.098	154135.734 11450.191	124421.914 12468.315	43469.434 2.783	167878.578 28299.416	43469.434 2.783	13.482 12471.098	167878.578 12471.098	43469.434 2.783	36256.879 2.783	36256.879 2.783	43469.434 2.783	CO2 H2	CO2 H2	
143432.969 0.984	143432.969 0.984	131681.281 0.904	38491.473	104942.266 0.985	93200.586 104941.445	93200.586 104941.445	0.818 0.984	143432.969 0.984	104942.266 0.985	93200.586 104942.266	104942.266 9846.616	93200.586 9846.616	N2 COS ZNO ZNS AL2O3	N2 COS ZNO ZNS AL2O3	

DSRP-b 12/20/97 pg.2

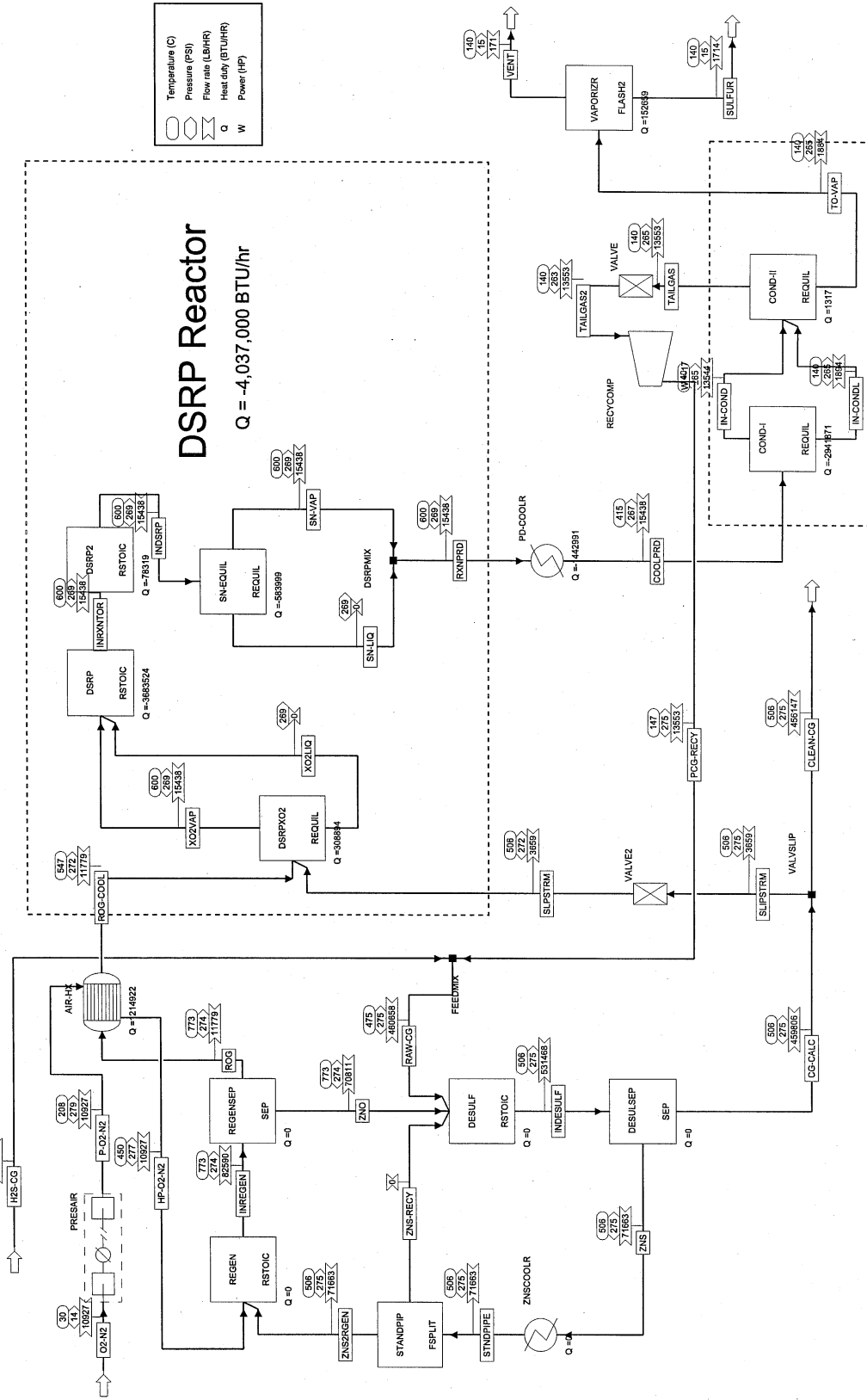
Display ALLS Units: Format: SOLI	TREAMS From DS To Phas	P-O2-N2 PRESAIR AIR-HX VAPOR	PCG-RECY RECYCOMP FEEDMIX DESULF VAPOR	RAW-CG FEEDMIX DESULF VAPOR	ROG REGENSEP AIR-HX VAPOR	ROG-COOL AIR-HX DSRPXO2 VAPOR	RXNRPD DSRPMIX PD-COOLR VAPOR	SLIPSTRM VALVE1 VALVE2 VAPOR	SLPSTRM VALVE1 VALVE2 VAPOR	SN-LIQ SN-EQUIL DSRPMIX MISSING	SN-VAP SN-EQUIL DSRPMIX VAPOR	STNDPIPE ZNSCOOL STANDPIP MISSING	SULFUR VAPORIZR LIQUID	Display ALLS Units: Format: SOLI	TREAMS From DS To Phas
Temperature [C]		207.7	147.5	418.7	810	586.3	600	647.5	647.5	647.5	600	640	140	Temperature [C]	
Pressure [PSI]		279.4	275	275	272.8	270.6	287.6	273.6	270.6	270.6	287.8	275	14.7	Pressure [PSI]	
Mass SFrac		0	0	0	0	0	0	0	0	0	0	0	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	0	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	-1.46E+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		0.006	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		0.006	H2	
O2			28299.416											O2	
N2			93200.566	104941.492	93200.566	93200.566	104942.266	11741.678	11741.678		104942.266		< 0.001	N2	
COS			0.984	0.984			0.985	0.081	0.081		0.985		0.985	COS	
ZNO													0.985	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	TREAMS
Units:	From DS To Phas	COND-II VALVE RECYCOMP VAPOR	COND-II VALVE RECYCOMP VAPOR	COND-II VAPORIZR LIQUID	VAPORIZR VAPOR	DSRPX02 DSRP MISSING	DSRPX02 DSRP VAPOR	DESULF MISSING	DESULF MISSING	DESULF MISSING	STANDPIP REGEN MISSING	From DS To Phas
Format: SOLI	Temperature [C]	140	139.9	140	140	600	600	810	810	640	640	Temperature [C]
Pressure	[PSI]	263.6	261	263.6	14.7		267.6	272.8	273.6	275	275	Pressure
Mass VFrac		1	0	1	0	0	1	0	0	0	0	Mass VFrac
Mass SFrac		0	0	0	0	0	0	1	1	1	1	Mass SFrac
*** ALL PHAS	ES ***											*** ALL PHAS
Mass Flow	[LB/HR]	163428.5	163428.5	21181.719	2136.962	0	184606.375	1191570	2668990	1467950	1201050	Mass Flow
Volume Flow	[CUFT/HR]	166691.297	168340.75	150.956	63551.742		398196.594	5717.191	12704.87	6987.678	5717.191	Volume Flow
Enthalpy	[BTU/HR]	-2.43E+08	-2.43E+08	-1.54E+07	-1.21E+07	0	-1.59E+08	-6.52E+09	-1.48E+10	-8.03E+09	-6.57E+08	Enthalpy
Density	[LB/CUFT]	0.98	0.971	140.317	0.034		0.464	208.419	210.076	210.076	210.076	Density
Mass Flow	[LB/HR]	194.606	194.606	0.499	0.497		37788.105					Mass Flow
O2S		239.99	239.99	0.353	0.352		5.113					O2S
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
CO		0.984	0.984	<0.001	<0.001		0.081					CO
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



High Pressure Condenser Q = -2,940,000 BTU/hr