Hot Gas Desulfurization With Sulfur Recovery[1]

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Abstract

Optimal development of advanced power processes requires that H_2S and other contaminants be removed at high temperature. Zinc-based sorbents for high temperature desulfurization have been developed to the stage of commercial-scale testing. However, problems associated with zinc sorbents including reduction to volatile $Zn(g)$, formation of $ZnSO_4$, and processing dilute SO_2 produced during regeneration have not been fully resolved.

This project examines an alternate process involving the direct production of elemental sulfur during regeneration. Current effort involves the sulfidation of CeO_2 to Ce_2O_2S and regeneration using SO_2 as the oxidant. Since CeO_2 is less efficient than ZnO for H_2S removal, a two-stage process may be necessary. Bulk H_2S removal using CeO_2 would be followed by a zinc sorbent polishing step. However, limited thermodynamic and literature data suggest that CeO_n (n<2) produced by reduction of CeO_2 may be capable of reaching the target H_2S concentration in a single stage.

Experimental effort has concentrated on the regeneration reaction which, in the 550 to 700 °C range, rapidly goes to completion with, at most, minimal side reactions. Elemental sulfur concentrations as large as 20% have been produced. About 800 °C is required for efficient sulfidation and H_2S concentration has been reduced to about 100 ppmv, the current detection limit of our analytical system. Ten cycle tests have shown negligible sorbent deterioration.

Sorbent durability is the key to process economics. Frequent sorbent replacement favors the zinc sorbent process with elemental sulfur recovery using the direct sulfur recovery process (DSRP). Increased sorbent durability should make the cerium sorbent process more economical.

Introduction

Desirable properties of a high temperature desulfurization sorbent include rapid reactivity with H_2S, ability to reduce the concentration of sulfur compounds to target levels ($\leq$ 20 ppmv), regenerability, and sufficient durability to maintain performance through many sulfidation-regeneration cycles. Recent attention has focused on zinc-based sorbents because of their favorable sulfidation properties. However, problems remain in the regeneration and durability areas. Because of the strong affinity between zinc and sulfur, regeneration of ZnS can only be accomplished by reaction with O_2 to reform ZnO and liberate SO_2. This highly exothermic reaction requires the use of dilute O_2 which produces dilute SO_2 and complicates the ultimate sulfur control problem. In addition, the presence of SO_2 and O_2 in the regeneration gas favors the formation of $ZnSO_4$, which is believed to be a major contributor to sorbent deterioration.

Direct production of elemental sulfur during sorbent regeneration, which is the primary objective of this study, would eliminate the SO_2 control problem and alleviate other problems associated with sorbent regeneration.

Project Description

This project began with a literature search which identified three concepts for the direct production of elemental sulfur during sorbent regeneration. This was followed by a thermodynamic analysis which evaluated a number of potential metal oxide sorbents using each regeneration concept (Lopez Ortiz, et al., 1995). Metal oxides having the strongest affinity for H_2S (e.g., ZnO) were least conducive to elemental sulfur production during regeneration. CeO_2, SnO_2, and, to a lesser degree, Fe_2O_3 provided a unique combination of H_2S removal capability and the potential for elemental sulfur production. The initial experimental effort examined the feasibility of FeS regeneration using the "partial oxidation" concept in which regeneration occurs in O_2-H_2O mixtures (Huang et al., 1996). While as much as 80% of the sulfur from FeS was liberated in elemental form, large amounts of excess steam were required and elemental sulfur concentrations in the product gas were low. Although feasible, this approach was judged to be commercially impractical.

At present we are examining the reduction, sulfidation, and regeneration reactions of CeO_2. At high temperature in a sufficiently strong reducing gas CeO_2 is reduced according to

$$CeO_2(s) + (2\text{-}n)\ H_2(g) ==> CeO_n(s) + (2\text{-}n)\ H_2O(g) \qquad \text{(R-1)}$$
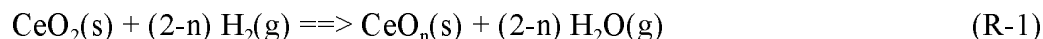
Equilibrium values of n as a function of temperature and partial pressure of O_2 have been reported by Bevan and Kordis (1964). The sulfidation reaction is

$$2\ CeO_n(s) + H_2S(g) + (2n\text{-}3)\ H_2(g) ==> Ce_2O_2S(s) + 2(n\text{-}1)\ H_2O(g) \qquad \text{(R-2)}$$
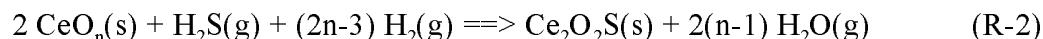
Although thermodynamic data for CeO_n is not available, the equilibrium H_2S removal potential increases as the value of n decreases. Ce_2O_2S may be regenerated with SO_2 according to the following reaction

$$Ce_2O_2S(s) + SO_2(g) ==> 2\, CeO_2(s) + S_2(g) \tag{R-3}$$
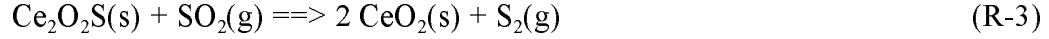
The sulfur product is represented as $S_2(g)$ for simplicity. In reality, a mixture of sulfur allotropes S_x with $1 \le x \le 8$ may be formed.

Thermodynamic Analysis

The thermodynamics of the H_2S-CeO_2 reaction are less favorable than for the H_2S-ZnO reaction. In a KRW gas containing 1% H_2S, ZnO in thermodynamically capable of removing about 99.6% of the H_2S at 800°C and 25 atm compared to about 65% H_2S removal by CeO_2. ZnO, however, cannot be used at this temperature because of its tendency to be reduced to Zn(g). No volatile products associated with CeO_2 are known. Further, as the reducing power of the coal gas increases, the desulfurization potential of CeO_2 also increases. In a Shell gas containing 1% H_2S at 800°C and 25 atm, CeO_2 is thermodynamically capable of removing about 97% of the H_2S; one would not consider using ZnO in this application.

Reduction of CeO_2 should begin at about 800°C in a Shell gas. Equilibrium calculations coupled with the results of Bevan and Kordis (1964) suggest that the composition should be CeO_2 at 650°C, $CeO_{1.9}$ at 800°C, and $CeO_{1.89}$ at 850°C. While the H_2S removal potential increases as the equilibrium value of n decreases, the free energy data for CeO_n is not available. The current status
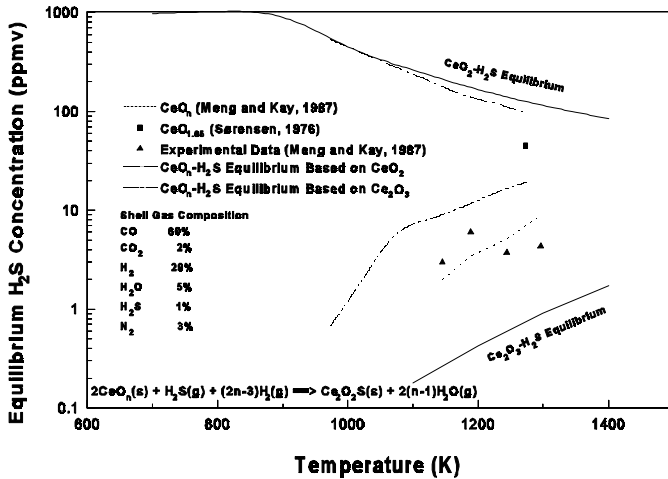

Figure 1. Sulfidation Thermodynamics

of equilibrium desulfurization of a Shell gas is summarized in Figure 1 where the equilibrium H_2S concentration (ppmv) is plotted versus temperature. The upper solid line represents CeO_2-H_2S equilibrium for which thermodynamic data is available. At 800°C, the equilibrium H_2S concentration is about 300 ppmv which corresponds to 97% H_2S removal. The lower solid line represents the H_2S-Ce_2O_3 ($CeO_{1.5}$) equilibrium for which thermodynamic data is also available. The equilibrium H_2S concentration at 800°C for this case is about 0.1 ppmv, well below target values. The dashed line and triangular data points represent equilibrium and experimental results reported by Meng and Kay (1987) using a gas composition similar to the Shell gas. While their method for estimating the equilibrium was not described, both the equilibrium and experimental results suggest that the H_2S concentration can be reduced to less than 10 ppmv in the 850 to 1000°C range. The single square point at 1000°C is based on single temperature thermodynamic data from Sørensen (1976) for $CeO_{1.85}$, and the calculated equilibrium H_2S concentration is 45 ppmv.

The two dot-dash curves were developed using approximate thermodynamic properties of CeO_n estimated from the properties of CeO_2 (upper curve) and Ce_2O_3 (lower curve). The upper curve shows only moderate improvement in H_2S removal compared to CeO_2, with improvement limited to temperatures above 800°C. The lower curve shows the potential for reducing H_2S levels to about 6 ppmv at 800°C. In spite of the differences in predicted H_2S equilibrium concentrations, the free energy values for CeO_n from the two approaches differ by only 1.5% at 800°C. This illustrates the sensitivity of equilibrium calculations to free energy values of the reacting species.
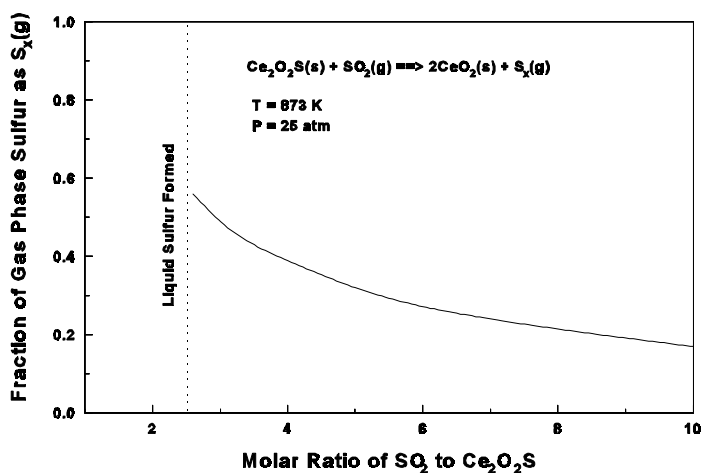

Figure 2. Regeneration Thermodynamics

In contrast to the uncertain sulfidation thermodynamics, the regeneration of Ce_2O_2S using SO_2 is highly favored. No significant by-products should be formed and high concentrations of elemental sulfur in the product gas may be produced. This is illustrated in Figure 2 where the equilibrium fraction of gas phase sulfur in elemental form is plotted versus the molar ratio of SO_2 to C_2O_2S at 873K and 25 atm. Complete regeneration is favored at all of the conditions shown; the process is limited not by incomplete regeneration, but by the condensation of elemental sulfur. While it should be possible to achieve complete regeneration using a SO_2 to Ce_2O_2S ratio of 3 to 1 with the product gas containing about 50% elemental sulfur, process analysis suggests that the concentration of elemental sulfur should be limited to about 15% because of heat transfer limitations associated with the exothermic regeneration reaction.

Experimental

The fixed-bed reactor shown in Figure 3 is being used to study the sulfidation and regeneration reactions. Sulfidation has used a mixture of H_2S, H_2, and N_2 while the regeneration atmosphere has consisted of SO_2 in N_2. Flow rates are controlled using mass flow controllers. The feed gas enters near the top of the stainless steel pressure vessel and flows downward over the sorbent held within a quartz insert. Product gases exit from the bottom and flow through a condenser and series of filters for removing elemental sulfur and then to a gas chromatograph for analysis .

The vapor pressures of H_2S and SO_2 have limited sulfidation and regeneration experiments to about 5 atm and 1 atm, respectively. In the future, gas mixtures will be purchased to enable higher pressures to be achieved. Condensation of elemental sulfur in tubing and filters has been a continuing problem. The current system of heat tracing lines and the arrangement of filters has evolved over time to reduce, but not eliminate, sulfur plugging. Sulfur deposited in the cooler sections of the
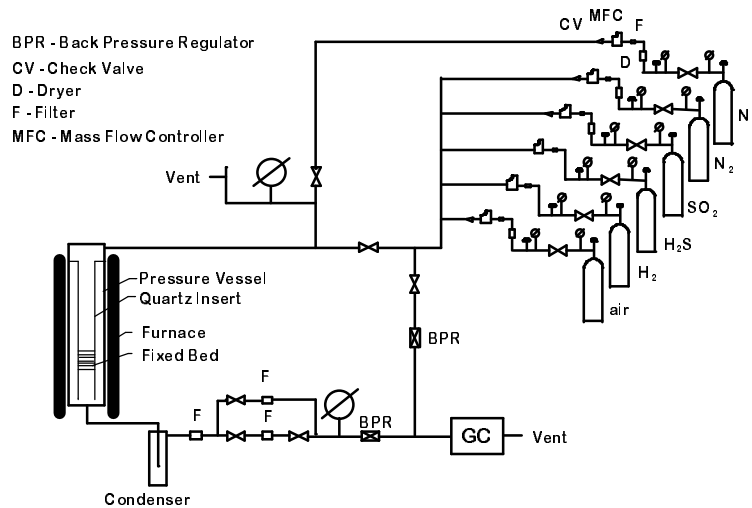
BPR - Back Pressure Regulator
CV - Check Valve
D - Dryer
F - Filter
MFC - Mass Flow Controller

Figure 3. Fixed Bed Reactor

reactor and downstream tubing during regeneration affects the apparent H_2S removal in subsequent sulfidation tests. The sulfur reacts with H_2 in the sulfidation gas which leads to abnormally high H_2S levels during the prebreakthough period. Cleaning procedures have been adopted to minimize this problem.

The gas chromatograph is equipped with a thermal conductivity detector (TCD) whose sensitivity for H_2S and SO_2 is limited to about 100 ppmv. A flame photometric detector (FPD) will be installed to extend the sensitivity to the levels required to determine the ultimate desulfurization capability of reduced CeO_n.

All tests have used high purity CeO_2 from Rhone-Poulenc physically mixed with Al_2O_3. Pure CeO_2 sinters to the point that it emerges from the reactor as a loosely bound, highly porous single cylinder. Sintering provides an open gas path around the circumference of the reactor insert which permits gas bypassing. No sintering has been observed with the CeO_2-Al_2O_3 mixtures. Free-flowing powders are removed at the end of the test. The CeO_2 is dry pressed into tablets which are then crushed and sieved. Most runs have used CeO_2 in the particle size range of 150 to 300 microns.

The range of sulfidation and regeneration conditions studied is summarized in Table 1. Six grams of CeO_2 with three grams of Al_2O_3 constitutes the maximum amount of sorbent which will fit within the isothermal zone of the reactor while containing sufficient Al_2O_3 to prevent sintering. The sulfidation rate was found to be effectively constant in the temperature range of 750 to 850°C, and appreciably smaller at 700°C. Sufidation feed gas rate was varied to reflect the mass of CeO_2 usedand to keep the duration of the sulfidation phase approximately constant. The same sulfidation gas composition has been used in all runs. Note that this gas composition contains no O_2 and is, therefore, more highly reducing than Shell gas. In the future, as greater attention is focused on prebreakthrough H_2S concentrations, it will be necessary to alter the gas composition to more accurately reflect the Shell gas composition. The regeneration reaction has received most of the attention to date since demonstrating elemental sulfur production is the primary objective of the study. The effects of temperature, gas composition (% SO_2), and gas flow rate have been examined. At temperatures of 450°C and below, the regeneration reaction is quite slow. Over the range of 500 to 700°C, the regeneration reaction is rapid and complete. 600°C has been chosen as the standard regeneration temperature. Regeneration runs have been limited to 1 atm because of the SO_2 vapor pressure. The SO_2 content of the regeneration gas has ranged from 1% to 20%, with the balance N_2. Of course, problems with elemental sulfur condensation and plugging become more severe as the SO_2 concentration is increased. Finally, the regeneration gas flow rate has been varied over the range of

Table 1. Reaction Conditions Used in Sulfidation and Regeneration Tests.

	Range	Standard
Sorbent		
CeO$_2$, g	3.0 - 6.0	6.0
Al$_2$O$_3$, g	0 - 6.0	3.0
CeO$_2$/Al$_2$O$_3$	0.5 - ∞	2
Sulfidation		
Temp., °C	700-850	800
Pressure, atm	5	5
GasComp.,%		
H$_2$S	1	1
H$_2$	10	10
N$_2$	89	89
Rate, sccm	200-400	400
Regeneration		
Temp., °C	350-700	600
Pressure, atm	1	1
Gas Comp., %		
SO$_2$	1-20	-
N$_2$	99-80	-
Rate, sccm	100-800	200

Figure 4. Sulfidation, Run Ce-17

100 to 800 sccm. The standard value of 200 sccm is not based on kinetic limitations, but upon the need to extend the duration of the regeneration test to permit sufficient data to be acquired to establish the shape of the breakthrough curve. Multicycle tests of ten and thirteen complete sulfidation and regeneration cycles have been completed to obtain preliminary information on sorbent durability. Standard sulfidation conditions were used in both multicycle runs. In the ten-cycle test the regeneration conditions were constant in each cycle while in the thirteen-cycle test regeneration gas composition and flow rate were varied.

Experimental Results

The H$_2$S breakthrough curves for the two sulfidation cycles of run Ce-17 are shown in Figure 4. The reactor system was extensively cleaned prior to both cycles to minimize residual sulfur contamination. Sulfidation was terminated when the H$_2$S content of the product gas reached about 0.1% (1000 ppmv). The prebreakthrough H$_2$S concentration ranged from 20 to 40 ppmv in cycle 1 and was approximately constant at 40 ppmv in cycle 2. While these numbers are, at best, approximate because of the TCD limitations, they show that concentrations below 100 ppmv
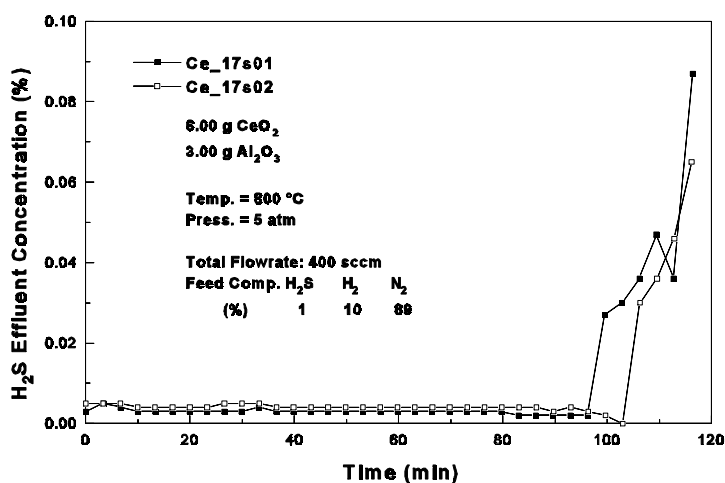
(>99% H_2S removal) can be achieved. The curves are effectively identical and there is no suggestion of sorbent deterioration. The prebreakthrough H_2S concentrations in sulfidation tests not preceded by extensive cleaning have been considerably larger than those shown in Figure 4 because of sulfur contamination.
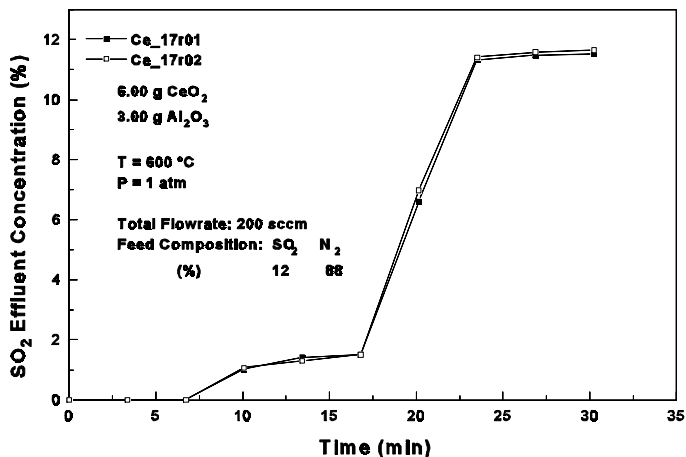

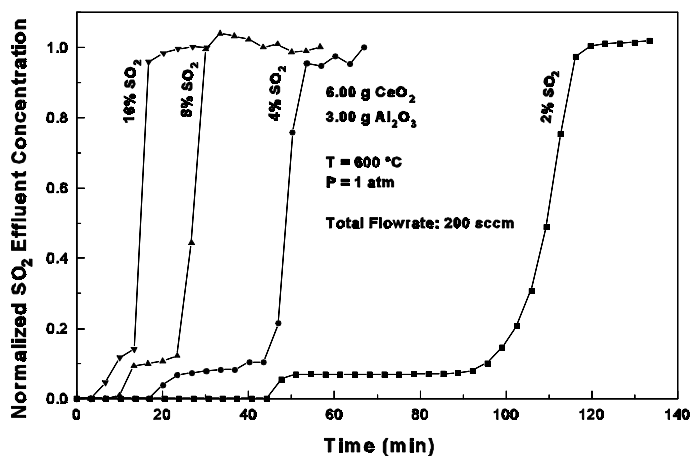
Figure 5. Regeneration, Run Ce-17

Figure 6. Effect of Sulfur Dioxide Concentration

SO_2 breakthrough curves from the two regeneration cycles of run Ce-17 are shown in Figure 5. No SO_2 was detected in the first three samples (t < ~ 7 min). The SO_2 concentration then increased to about 2% over the next three cycles, and closely approached the 12% concentration of the feed gas by the eighth sample (t ~ 22 min). Results of the two cycles were effectively equal and elemental sulfur concentration (considered as S_2) in the product gas was equal to or exceeded 10% for the first 17 minutes. The shape of the regeneration breakthrough curves in Figure 5 is typical, as illustrated in Figure 6 where the effect of regeneration gas composition (from 2% to 16% SO_2) is shown. The product SO_2 concentration is normalized so that the final concentration should approach 1.0 in all cases. All SO_2 concentrations were zero initially, rose to a plateau level between 0.07 and 0.15, and then increased rapidly to final normalized concentration of 1.0. Analytical limitations associated with large SO_2 concentration are illustrated by the breakthrough curve corresponding to 16% SO_2. Regeneration was effectively complete after only 6 samples (~17 min), and the product concentration increased from 0.14 to 0.96 between samples 5 and 6. With higher SO_2 concentrations the characteristics of the breakthrough curve would be lost completely.

Preliminary durability results are illustrated in Figures 7 and 8, the sulfidation and regeneration breakthrough curves from ten-cycle test Ce-16. Figure 7 contains the H_2S breakthrough curves from cycles 2 through 10 of this test. An error in mass flow controller settings was made in cycle 1 and
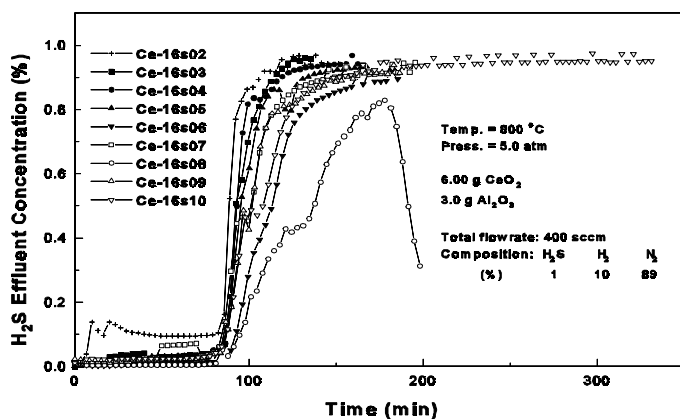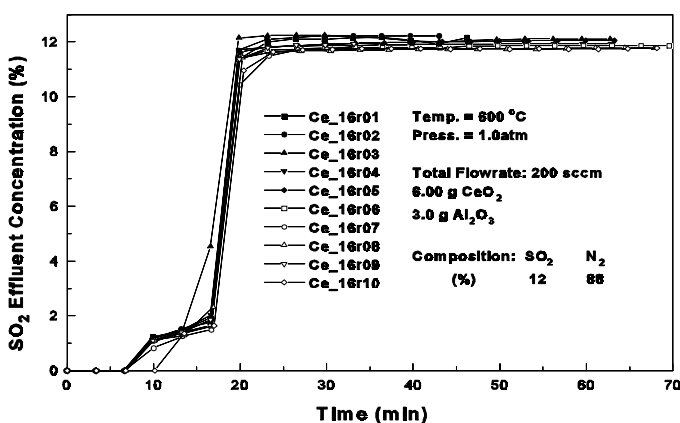
Figure 7. Sulfidation, Run Ce-16

Figure 8. Regeneration, Run Ce-16

the results are not comparable. Prebreakthrough H₂S concentrations were higher than those shown in Figure 4, but the reactor system was, at best, only minimally cleaned following each regeneration cycle.

The breakthrough time, i.e., the time corresponding to 0.05% H_2S in the product ranged from a low of 79.5 minutes in Ce-16s04 to a maximum of 84.3 minutes in Ce-16s05. Due to poor cleaning, results from Ce-16s02 were above the 0.05% H_2S limit from the beginning and were omitted from the breakthrough time summary. Also, the unexplained upset in Ce-16s07 in the 50 to 70 minute period was ignored in determining breakthrough time. With the exception of Ce-16s08, all sulfidation cycles went to completion in the same general time period. The H_2S mass flow controller in Ce-16s08 malfunctioned after about 175 minutes, which may also have been responsible for the earlier differences in the H_2S breakthrough curve for that cycle. All regeneration cycles of run Ce-16 used the same reaction conditions and SO₂ breakthrough curves are shown in Figure 8. With the exception of two samples--one at 17 minutes in Ce-16r03 and the other at 10 minutes in Ce-16r10--the results were effectively identical. The first measurable quantity of SO₂, about 1%, was detected after 10 minutes, and by 23 minutes regeneration was effectively complete. The final SO₂ content of the product gas ranged from 11.8 to 12.2%.

Sulfur material balances, expressed as percent of stoichiometric sulfur removed during sulfidation and liberated during regeneration, are presented in Figure 9. Results from sulfidation cycles Ce-16s01 and Ce-16s08 are omitted because of the previously described problems. In the eight sulfidation cycles the sulfur balance ranged from a minimum of 75% of stoichiometric in Ce-16s02 to a maximum of 96.8% in Ce-16s10. Regeneration results ranged from a minimum of 86.0% of stoichiometric in Ce-16r03 to 92.6% in Ce-16r10. The maximum difference in sulfur balance within a single cycle occurred in cycle 2 where sulfur removal and liberation were 75.0% and 90.4% of stoichiometric, respectively. However, sulfur removal in the eight sulfidation cycles averaged 87.2% of stoichiometric (88.9% if the low value associated with Ce-16s02 is omitted), while sulfur liberated in the ten regeneration cycles averaged 91.8% of stoichiometric. The low sulfur removal in
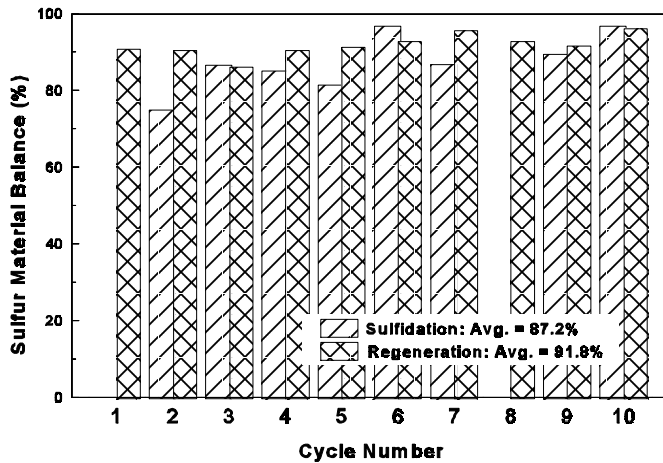
Figure 9. Sulfur Material Balance, Run Ce-16

Ce-16s02 may be due to the fact that the reactor was only minimally cleaned following regeneration cycle Ce-16r01. This is shown by the large prebreakthrough H_2S concentration in Figure 7.

The results of this ten-cycle run are judged to be quite favorable. The constancy of the slopes of the sulfidation and regeneration breakthrough curves, the small variation in breakthrough times, and the good sulfur material balance closure all suggest little, if any, sorbent deterioration.

Process Analysis

The process flow diagram shown in simplified form in Figure 10 was developed on the basis of the thermodynamic analysis and experimental results. Two-stage desulfurization is proposed with the bulk of the H_2S removed by cerium sorbent in the primary sorber followed by a polishing stage using a zinc-based sorbent to reduce H_2S concentration to IGCC target levels. SO_2 regeneration of Ce_2O_2S in the primary regenerator produces elemental sulfur directly. Zinc sorbent is regenerated using O_2, and SO_2 is recycled to the gasifier. The recycle stream and the gasifier are shown by dashed lines to indicate that they were not included in the material and energy balance calculations.
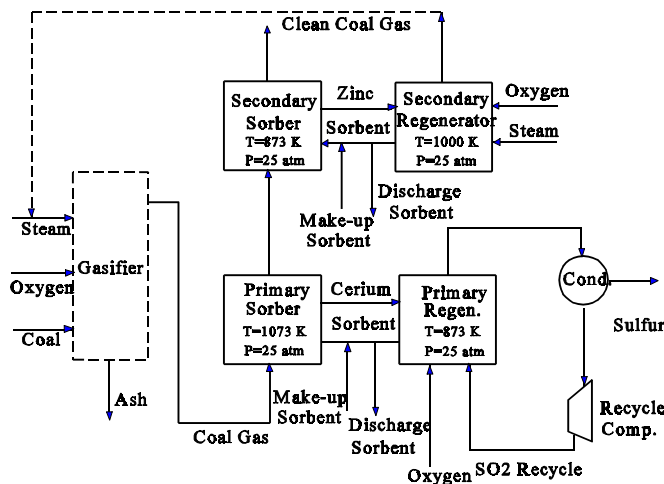
A variation of the two-stage process in which it was arbitrarily assumed that H_2S target levels could be achieved using cerium sorbent in a single stage was also considered. The justification for this option is based on the earlier thermodynamic discussion but the feasibility has not yet been proven. The process is similar to that shown in Figure 10 except that the secondary sorber and regenerator are omitted and no recycle stream to the gasifier is needed.

Figure 10. Two-Stage Cerium Desulfurization with SO_2 Regeneration

The cerium-based processes have been compared to a traditional single-stage process using a zinc sorbent with elemental sulfur recovery using the DSRP process (McMichael and Gangwal, 1997). The simplified diagram of this process is shown in Figure 11. H_2S concentration is reduced to IGCC levels in the sorber; the sulfided sorbent is regenerated in the traditional manner to liberate SO_2 which is subsequently reduced to elemental sulfur by diverting a slip stream of the coal gas to the DSRP reactor.
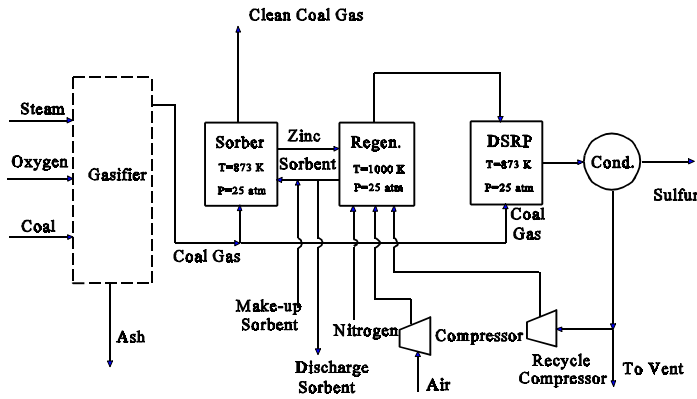

Figure 11. Single-Stage Zinc Desulfurization with DSRP

Nominal operating conditions for the reactors are shown on the figures and a summary of the overall process material balances is presented in Table 2. In each case the feed is 17,000 lbmol/hr of Shell gas containing 5500 lb/hr of sulfur as H_2S. The flow rate was chosen to match that used by Buchanan et al. (1994) in their evaluation of IGCC desulfurization options. There is no reduction in the flow rate of clean coal gas using either of the cerium sorbent options while 500 lbmol/hr of coal gas is lost to the DSRP reactor in the zinc sorbent process. The H_2S concentration in the clean coal gas is 10 ppmv in each case; the quantities of elemental sulfur produced are almost the same. There are no direct discharges of sulfur to the environment from either of the cerium processes but approximately 5 lb/hr of sulfur is present in the purge gas from the zinc sorbent process.

Table 2. Summary of Process Material Balances.

	Cerium Sorbent		Zinc Sorbent
	Two-Stage	Single-Stage	Single-Stage
Coal Gas Feed, lbmol/hr	17,000	17,000	17,000
Sulfur Content, lb/hr	5500	5500	5500
Coal Gas Product, lbmol/hr	17,000	17,000	16,500
Sulfur Content, lb/hr	6	6	6
H_2S Concentration, ppmv	10	10	10
Elemental Sulfur Product, lb/hr	5494	5494	5489
Sulfur Discharge (SO_2), lb/hr	-	-	5

Major equipment items for each option were sized and capital requirements estimated using information from the literature and informal vendor quotes. Operating costs were estimated on the basis of raw material and utility requirements with appropriate credits taken for excess steam and elemental sulfur. The final cost comparison was based on the 10-year levelized cost, and the sensitivity of the levelized cost to selected cost items was determined.

Sorbent cost, which is a function of the sorbent unit cost and replacement rate, is the most important single cost item. Levelized costs for the two-stage cerium process and single-stage zinc process are compared in Figure 12. Unit costs for both zinc and cerium sorbents of $8/lb, $5/lb, and $3/lb are considered, and the sorbent replacement rate, expressed as percent of the sorbent circulation rate between the fluidized-bed sorbers and regenerators, is varied between 1% (average sorbent life of 100 cycles) and 0% (infinite sorbent life). The three solid lines represent the estimated cost of the two-stage cerium process while the dashed lines represent single-stage desulfurization with zinc sorbent. Large sorbent replacement rates (small sorbent lifetime) favor the zinc sorbent process. For example, if the unit cost of both sorbents is $5/lb, the breakeven sorbent replacement rate is about 0.25% (400 cycles average lifetime) and the estimated annual levelized cost is about $8.8 million. Larger replacement rates increase the process cost and favor the zinc sorbent process while smaller replacement rates reduce process cost and favor the two-stage cerium sorbent process. The annual levelized cost becomes independent of the sorbent unit cost when sorbent replacement is not necessary, i.e., for infinite sorbent life. For the zinc sorbent process all lines approach an annual levelized cost of about $5 million while the lines for the cerium sorbent process converge to -$2.2 million. In other words, the cerium process has the possibility of showing a profit if the sorbent is sufficiently durable.
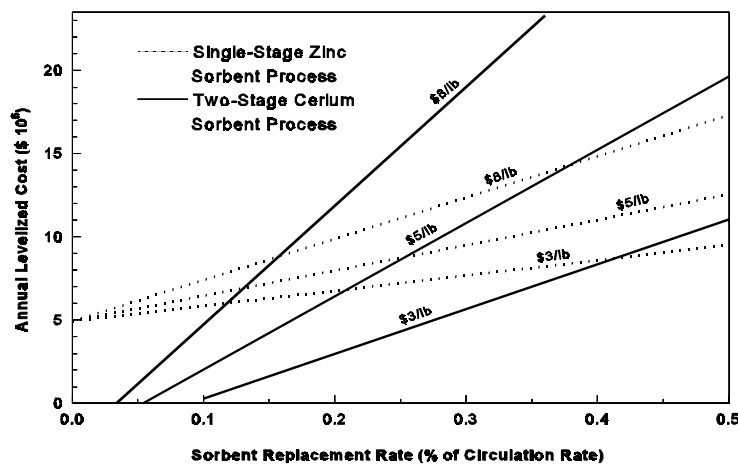

Figure 12. Annual Levelized Cost Comparison

The third process, which assumes that single-stage cerium desulfurization will be satisfactory for reducing the H_2S concentration to IGCC levels in a single step, results in additional process simplication and cost savings. The secondary sorber and regenerator along with the zinc sorbent used in those units will not be needed. Steam and oxygen to the secondary regenerator are not needed. The value of the clean coal gas will be increased since it is available to the power generation section at higher temperature. Finally, the desulfurization-sulfur recovery section will be uncoupled from the gasifier since recycle of the secondary regenerator product gas is not required. The increased value of the higher temperature coal gas and elimination of the secondary regenerator recycle are not reflected in the economic evaluation, which includes only the capital cost reduction, elimination of the zinc sorbent, and relatively minor changes in utility requirements and steam by-

product. The estimated savings compared to the two-stage cerium process a levelized cost basis is $350,000 per year. The significance of this saving depends on the total process cost. If the cost of the cerium sorbent is $5/lb and the required replacement rate is 0.5% of the circulation rate, the $350,000 savings is only a small percentage of the estimated annual levelized cost of the two-stage cerium process of $20 million. The zinc sorbent process at an annual levelized cost of about $13 million would still be more economical, although it is doubtful that either process would be commercially attractive. However, if the sorbent replacement rate is only 0.1% of the circulation rate and the unit sorbent cost is $5/lb, the $350,000 saving is more than 10% of the estimated $2.2 million annual levelized cost of the two-stage cerium process. Both of the cerium options are more economical than the zinc process at these conditions.

Summary and Future Activities

The regeneration of Ce_2O_2S with SO_2 proceeds rapidly at 600°C with stoichiometric production of elemental sulfur concentration of as much as 20%, above the optimum level of about 15% identified in the process analysis. Temperatures of approximately 800°C are required for efficient sulfidation, and H_2S concentrations have been reduced from 10,000 ppmv to near the detection limit of the TCD, about 100 ppmv. No evidence of CeO_2 sorbent deterioration was seen in a ten-cycle test.

The major remaining objective is to determine the ultimate desulfurization potential of reduced CeO_n. This will require that the reactor system be modified to eliminate sulfur contamination from previous regeneration tests, and installation of a flame photometric detector (FPD) in the chromatograph to permit accurate analysis of H_2S concentrations in the low ppm range.

In addition, the effects of reaction parameters such as temperature, pressure, and gas flow rate and composition will be more fully explored. CeO_2 from at least three sources will be tested. Because of the economic importance of sorbent durability, additional multicycle tests will be conducted. Sufficient information to guide future larger-scale studies in fluidized-bed reactors will be gathered.

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