A Calcium Oxide Sorbent Process for Bulk Separation of Carbon Dioxide

Authors:
Douglas P. Harrison
Chun Han
George Lee

Contractor:
Louisiana State University
Department of Chemical Engineering
Baton Rouge, Louisiana 70803

Contract Number:
DE-AC21-89MC26366

Conference Title:
Advanced Coal-Fired Power Systems '95 Review Meeting

Conference Location:
Morgantown, West Virginia

Conference Dates:
June 27-29, 1995

Conference Sponsor:
U.S. Department of Energy, Morgantown Energy Technology Center (METC)
DISCLAIMER

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

This report has been reproduced directly from the best available copy.

Available to DOE and DOE contractors from the Office of Scientific and Technical Information, 175 Oak Ridge Turnpike, Oak Ridge, TN 37831; prices available at (615) 576-8401.

Available to the public from the National Technical Information Service, U.S. Department of Commerce, 5285 Port Royal Road, Springfield, VA 22161; phone orders accepted at (703) 487-4650.
OBJECTIVES

This project began as an experimental investigation of the removal of CO₂ from coal-derived gas at moderate to high temperature using a calcium oxide sorbent. The project evolved into a study of the simultaneous water-gas shift reaction and CO₂ removal to prove the technical feasibility of an alternate process for the production of H₂ from coal-derived gas. The important stoichiometric reactions are:

\[ \text{CO(g)} + \text{H}_2\text{O(g)} \leftrightarrow \text{CO}_2\text{(g)} + \text{H}_2\text{(g)} \]  \hspace{1cm} (1)

\[ \text{CaO(s)} + \text{CO}_2\text{(g)} \leftrightarrow \text{CaCO}_3\text{(s)} \]  \hspace{1cm} (2)

Favorable results from the experimental tests prompted METC to undertake an Aspen simulation study to carry out the material and energy balance calculations on a conceptual process based upon the above chemistry. This paper summarizes the most relevant of the experimental data and the Aspen simulation results.

BACKGROUND INFORMATION

Hydrogen is key raw material for many of the most important chemical products such as ammonia and methanol (Shreve and Brink, 1977). It is also becoming increasingly important in
petroleum refining as the rising demand for gasoline coupled with the use of heavier feedstocks result in increased reliance on the various hydropressing operations (Abrardo and Khurana, 1995). The majority of hydrogen is produced by the steam reforming or partial oxidation of gaseous or liquid hydrocarbons. In the future, coal gasification may become an important raw material for H₂ manufacture. The primary products from steam reforming, partial oxidation, and coal gasification - CO, CO₂, H₂ and H₂O - are the same.

Synthesis gas is normally upgraded to high purity hydrogen using the water-gas shift reaction followed by CO₂ removal. Multiple catalytic reaction steps are required to achieve the required fractional conversion of CO and the maximum H₂ production. CO₂ removal is accomplished using wet scrubbing or pressure swing adsorption. Extensive heat exchange networks are required since various steps in the process operate at different temperature.

In contrast, the combined shift-carbonation process accomplishes shift reaction and CO₂ separation in a single processing vessel. No shift catalyst is required because of the higher operating temperature, and heat exchange requirements are reduced because the alternate process operates under more nearly isothermal conditions.

PROJECT DESCRIPTION

A laboratory-scale fixed-bed reactor having a capacity of 10 to 15 grams of calcium-based CO₂ acceptor was used in the experimental study. Dolomite was found to be superior to limestone as a sorbent precursor and all experimental results and Aspen simulations presented in this paper are based upon dolomite. Gaseous feed components -- CO, CO₂, H₂, and N₂ -- were obtained from high pressure cylinders with flow rate controlled using high pressure mass flow controllers. Water was added using a high pressure syringe pump. The water was vaporized as it mixed with the permanent gases by heat tracing the feed lines, and the combined feed was preheated to reaction temperature prior to contacting the sorbent. Reactor feed or product was sampled periodically and analyzed using gas chromatography. A complete description of the reactor and analytical system has been published (Harrison et al., 1993).

Reaction parameters studied included calcination and carbonation temperature and pressure, feed gas composition, reactor space velocity, and sorbent composition and properties. Only a brief summary of the most relevant experimental test results is presented in this paper; more detailed results are available elsewhere (Han and Harrison, 1994; Harrison and Han, 1994; Harrison, et al. 1994).

The final phases of the experimental study were carried out concurrently with the Aspen simulation work performed at METC. Key assumptions involved in the Aspen simulation were that the rates of the shift and carbonation reactions were sufficiently fast to permit equilibrium to be achieved, that 80% conversion of CaO to CaCO₃ was possible in the carbonation step, and that the average sorbent lifetime was 20 cycles. Each of these assumptions was justified by the experimental results.

SUMMARY OF EXPERIMENTAL RESULTS

A typical reactor response curve showing concentrations of H₂, CO₂, and CO in the reactor product gas as a function of dimensionless time, t*, is shown in Figure 1. During the early stages of the reaction, referred to as the prebreakthrough period, the CO and CO₂ concentrations were approximately constant in the range of 80 and 300 ppmv (dry basis), respectively, and the H₂ concentration was a maximum at about 15% (vol). Both the shift and carbonation reactions were fully effective during this period, and these concentrations correspond to 0.998 fractional removal of total carbon oxides. Equilibrium concentrations at these reaction conditions are 20
ppm CO and 110 ppm CO₂, which correspond to 0.999 equilibrium carbon oxide removal. CO and CO₂ concentrations began to increase rapidly at t* ~ 0.7 and continued to climb until a second steady-state was reached at t* ~ 1.1. During this active breakthrough period the H₂ content decreased from about 15% to 9.4% (dry basis). The beginning of breakthrough corresponded to the time at which the leading edge of the carbonation reaction front reached the exit of the packed bed. During the postbreakthrough period, effectively all of the CaO had been converted to CaCO₃, and the water-gas shift was the only reaction occurring to any significant extent.

Fractional carbon oxide removal, FCOX, and fractional conversion of CaO, X*, as a function of t* for this test are shown in Figure 2. Both parameters were calculated directly from the measured concentration-time data of Figure 1. The prebreakthrough value of FCOX remained above 0.99 for t* ≤ 0.85. The definition of the dimensionless time is such that t* = X* whenever the fractional removal of carbon oxides is complete, i.e., as long as FCOX = 1. Thus, X* was effectively equal to t* during the prebreakthrough period. The fractional conversion of CaO increased linearly with time until breakthrough began; thereafter the rate of conversion decreased quickly and reached a final value of X* = 0.93.

The results illustrated in Figures 1 and 2 were typical. FCOX values in excess of 0.99 and final X* values greater than 0.9 were achieved over a wide range of shift-carbonation temperatures, space velocities, and feed gas compositions. Single-cycle tests proved the feasibility of combining the shift and carbonation reactions to produce a product gas rich in hydrogen with carbon oxide removals closely approaching theoretical equilibrium limits. The
second major question addressed in the experimental effort was that of sorbent multicycle durability. Although dolomite is widely available and inexpensive, it must maintain the favorable reaction properties through a number of calcination-carbonation cycles if the process is to be commercially attractive. A number of multicycle tests consisting of five, and in two cases eleven, complete cycles were conducted to obtain preliminary information on dolomite durability.

Figure 3 compares the CO and CO₂ concentrations and the fractional removal of total carbon oxides during the prebreakthrough period of an eleven-cycle and a five-cycle test at the same reaction conditions. CO concentrations were typically in the 20 to 30 ppmv range and CO₂ concentrations, although somewhat more variable, were generally in the 180 to 250 ppmv range. These concentrations correspond to fractional removal of total carbon oxides of either 0.996 or 0.997 in each of the 16 cycles.

Although prebreakthrough concentrations did not increase with increasing number of cycles, sorbent deterioration did occur as shown by the decrease in both the duration of the prebreakthrough period and the slope of the breakthrough curve. These factors are illustrated in Figure 4 where CO₂ breakthrough curves for the first, fifth, and eleventh cycles are compared. The duration of the prebreakthrough period decreased from $t^* \sim 0.8$ in the first cycle to $t^* \sim 0.6$ in the fifth cycle to $t^* \sim 0.25$ in the eleventh cycle. It was on the basis of such multicycle tests that the average sorbent lifetime of twenty calcination-carbonation cycles used in the process analysis was established.
Figure 3. Multicycle Results During the Prebreakthrough Period

Figure 4. Decreased Duration of the Prebreakthrough Period
PROCESS SYNTHESIS AND ANALYSIS

METC and LSU researchers collaborated to define the processing steps needed to produce H₂ using the combined shift-carbonation process. METC personnel then used Aspen simulation to carry out material and energy balance calculations.

The Process

A block flow diagram of the combined shift-carbonation process is shown in Figure 5. A sulfur-free coal gas at 1925°F and 412 psia whose flow rate and composition are shown in the first column of Table 1 passes through a heat recovery steam generator where 5x10⁴ lb/hr of 400 psia, 1000°F steam is generated by cooling the coal gas to 1100°F. The coal gas stream is then split with approximately one-half fed to the shift-carbonation reactor and the remainder used as fuel to supply heat required for sorbent regeneration. 1.6x10⁴ lb/hr of the 400 psia, 1000°F steam is added to the coal gas to produce a 1.3 to 1 molar ratio of steam to carbon monoxide in the reactor feed (see column two of Table 1). This ratio is considerably less than used in a traditional shift reactor. The remainder of the steam is available for power generation or "sale."

The combined coal gas and steam are fed to the shift-carbonation reactor where they contact calcined dolomite. The product gas flow rate and composition, shown in the third column of Table 1, are based upon combined shift-carbonation reaction equilibrium at 1022°F and 340 psia. As shown previously, the experimental results suggest that equilibrium should be closely approached at these conditions. CO and CO₂ concentrations in the reactor product gas stream are shown as zero in Table 1 although the actual equilibrium concentrations are approximately 25 ppm CO₂ and

Figure 5. The Combined Shift-Carbonation Process
Table 1. Flow Rate and Composition of Gas Streams in the Shift Carbonation Process

<table>
<thead>
<tr>
<th>Stream Description</th>
<th>Desulfurized Coal Gas</th>
<th>Shift-Carbonation Reactor Feed</th>
<th>Shift-Carbonation Reactor Product</th>
<th>Hydrogen Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temp., °F</td>
<td>1925</td>
<td>1050</td>
<td>1020</td>
<td>59</td>
</tr>
<tr>
<td>Pressure, psia</td>
<td>412</td>
<td>400</td>
<td>340</td>
<td>340</td>
</tr>
<tr>
<td>Flow Rate, lb mol/hr</td>
<td>1.00x10^4</td>
<td>5.81x10^3</td>
<td>3.69x10^3</td>
<td>3.08x10^3</td>
</tr>
<tr>
<td></td>
<td>1.99x10^5</td>
<td>1.14x10^5</td>
<td>2.18x10^4</td>
<td>1.09x10^4</td>
</tr>
<tr>
<td>Composition, mol fraction</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>N₂</td>
<td>0.012</td>
<td>0.010</td>
<td>0.016</td>
<td>0.019</td>
</tr>
<tr>
<td>H₂</td>
<td>0.251</td>
<td>0.213</td>
<td>0.759</td>
<td>0.909</td>
</tr>
<tr>
<td>CO</td>
<td>0.336</td>
<td>0.284</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.093</td>
<td>0.079</td>
<td>0.000</td>
<td>0.000</td>
</tr>
<tr>
<td>H₂O</td>
<td>0.261</td>
<td>0.374</td>
<td>0.164</td>
<td>0.000</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.042</td>
<td>0.036</td>
<td>0.056</td>
<td>0.067</td>
</tr>
<tr>
<td>NH₃</td>
<td>0.003</td>
<td>0.003</td>
<td>0.004</td>
<td>0.005</td>
</tr>
</tbody>
</table>

30 ppm CO. The mol fraction H₂ in the product stream is 0.759, with the impurities consisting primarily of CH₄, N₂, and H₂O having mol fractions of 0.056, 0.016, and 0.164, respectively. The remaining H₂O would be removed by cooling and condensation (not shown in Figure 4) to produce a final product containing about 91% H₂ (column 4 in Table 1). CH₄ and N₂ entered with the coal gas and were assumed to be inert in the shift-carbonation simulation. In order to produce a product containing greater than 99% H₂, the product from the gasifier would have to be essentially CH₄- and N₂-free.

Both the shift and carbonation reactions are exothermic, and the energy liberated by the reaction is used to generate an additional 2.3x10^5 lb/hr of 400 psia, 1000°F steam. The total quantity of excess steam produced in the process is about 2.6 x 10^5 lb/hr.

The feed rate of fully calcined dolomite to the shift-carbonation reactor of 2.6x10^5 lb/hr is sufficient to provide 20% excess over the stoichiometric quantity required for complete reaction with the CO and CO₂ present in the coal gas. Spent sorbent is transferred to the regenerator for decomposition of CaCO₃. Approximately 5% (1.8x10^4 lb/hr) of the spent sorbent is purged and replaced by an equivalent quantity of make-up dolomite to maintain the required sorbent reactivity.

Regeneration, which takes place at 1650°F and 19 psia, is assumed to be complete, in agreement with experimental results. Heat required for the endothermic regeneration reaction is supplied by the combustion of a portion of the coal gas with preheated air. Because of the large quantity of energy required, about 50% of the coal gas must be used for this purpose, and the production rate of H₂ is reduced proportionally. The total energy required by the endothermic regeneration reaction is almost balanced by the exothermic energy liberated by the shift and carbonation reactions. However, since calcination requires higher temperature, energy exchange is not possible.

Certain heat exchangers, boiler feed water pumps, and a regeneration air blower are not shown in Figure 5. However, the heat duties of
the exchangers were included when calculating the total steam production rate. The total power required by the pumps and blower was estimated to be $1.7 \times 10^7 \text{ Btu/hr}$. A summary of overall process inputs and outputs, including the pumps and blower, is presented in Table 2. The important input streams are the desulfurized coal gas at a rate of $1.0 \times 10^4 \text{ lb mol/hr}$ ($1.99 \times 10^5 \text{ lb/hr}$), and $2.6 \times 10^5 \text{ lb/hr}$ of make-up dolomite. From these feeds, $3.1 \times 10^3 \text{ lb mol/hr}$ ($1.09 \times 10^4 \text{ lb/hr}$) of 91% $\text{H}_2$ and $2.63 \times 10^3 \text{ lb/hr}$ of 400 psia, 1000°F steam are produced. Complete results of the material and energy balance calculations for all process streams are available in the final report (Harrison et al., 1994).

Comparison With Other Processes for $\text{H}_2$ Production

As part of a recent DOE-sponsored study on the production of $\text{H}_2$ from coal gas using a high temperature membrane process, Jungerhans (1992) carried out similar material and energy balance calculations to compare the high temperature membrane process with a conventional process using multiple catalytic shift reactors and pressure swing adsorption (PSA) for $\text{H}_2$ purification. The coal gas feed rate used by Jungerhans was the same as used in this study, and the composition of the coal gases were similar as shown in Table 3. Differences in the

![Table 2. Summary of Process Input and Output Streams](image)

-662-
composition are important, however. Because of the reduced \( \text{H}_2 \) and \( \text{CO} \) content of the coal gas used in the present study, the maximum production rate of \( \text{H}_2 \) (assuming complete conversion of \( \text{CO} \) via the shift reaction and complete \( \text{H}_2 \) recovery) is 13% less. Actual \( \text{H}_2 \) production rates will, of course, be significantly lower due to incomplete \( \text{CO} \) conversion, losses during purification, and the necessity to divert a portion of the coal gas for use as fuel. \( \text{CH}_4 \) and \( \text{N}_2 \) impurities were assumed to be totally removed in the PSA unit of the conventional process, and the membrane was considered to be impermeable to \( \text{CH}_4 \) by Jungerhans.

In spite of these differences, it is instructive to compare the \( \text{H}_2 \) product from the combined shift-carbonation process to the product from the two processes considered by Jungerhans (see Table 4). The quantity of \( \text{H}_2 \) produced in the conventional shift and membrane processes is

### Table 3. Comparison of the Coal Gas Feed Composition Evaluated by Jungerhans and This Study

<table>
<thead>
<tr>
<th></th>
<th>Jungerhans (1992)</th>
<th>This Study</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow Rate, lb mol/hr</td>
<td>1.0x10^4</td>
<td>1.0x10^4</td>
</tr>
<tr>
<td>Mol Fraction</td>
<td></td>
<td></td>
</tr>
<tr>
<td>( \text{H}_2 )</td>
<td>0.289</td>
<td>0.251</td>
</tr>
<tr>
<td>( \text{CH}_4 )</td>
<td>0.050</td>
<td>0.042</td>
</tr>
<tr>
<td>( \text{CO} )</td>
<td>0.386</td>
<td>0.337</td>
</tr>
<tr>
<td>( \text{CO}_2 )</td>
<td>0.107</td>
<td>0.093</td>
</tr>
<tr>
<td>( \text{H}_2\text{S} )</td>
<td>0.001</td>
<td>0.000</td>
</tr>
<tr>
<td>( \text{H}_2\text{O} )</td>
<td>0.150</td>
<td>0.261</td>
</tr>
<tr>
<td>( \text{N}_2 )</td>
<td>0.017</td>
<td>0.012</td>
</tr>
<tr>
<td>( \text{NH}_3 )</td>
<td>0.000</td>
<td>0.003</td>
</tr>
</tbody>
</table>

### Table 4. Comparison of the Products from Various Processes for the Production of Hydrogen

<table>
<thead>
<tr>
<th></th>
<th>Membrane Separation</th>
<th>Conventional Shift</th>
<th>Combined Shift-Carbonation</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \text{H}_2 ) Product \lb mol/hr</td>
<td>4637</td>
<td>4490</td>
<td>3080</td>
</tr>
<tr>
<td>% ( \text{H}_2 )</td>
<td>99.9</td>
<td>100</td>
<td>91.0</td>
</tr>
<tr>
<td>% of Theoretical ( \text{H}_2 )</td>
<td>68.6</td>
<td>66.5</td>
<td>47.7</td>
</tr>
</tbody>
</table>
larger, both on an absolute basis and as a percentage of the theoretical maximum production rate. \( H_2 \) is lost in the membrane process due to incomplete membrane permeation. The nonpermeate stream is used for power generation in a combustion turbine. The small impurity is primarily \( CH_4 \) formed by the methanation of CO and CO\(_2\) which permeate the membrane. In the conventional shift process, \( H_2 \) losses are associated with diversion of 27% of the coal gas for power generation and from the off-gas of the PSA unit. None of the heavier components escape the PSA separator so that the \( H_2 \) content of the product is essentially 100%. The choice of PSA over amine-based wet scrubbing for \( H_2 \) purification was dictated by the high \( CH_4 \) concentration in the coal gas. \( CH_4 \) would not be removed in the wet scrubber and the \( CH_4 \) content of the product would be comparable to that produced in the shift-carbonation process.

The lower \( H_2 \) production rate in the shift-carbonation process is caused by the need to divert approximately 50% of the coal gas to supply energy for spent sorbent regeneration. The lower \( H_2 \) content is due, as previously discussed, to the \( CH_4 \) and \( N_2 \) content of the coal gas. This problem would not exist if the coal gas was free of \( CH_4 \) and \( N_2 \), in which case the \( H_2 \) purity would be comparable to that achieved with the other processes.

**Process Improvements**

A number of process improvements and/or modifications would be required for the shift-carbonation process to become economically competitive. Three are briefly discussed in the following.

The \( H_2 \) product purity is an obvious problem. Increasing the \( H_2 \) content could be accomplished by adding an additional purification step, probably a PSA unit. This option would, of course, further reduce the \( H_2 \) production rate; \( H_2 \) losses in a PSA separator typically are about 20% (Ruthven, 1984). Alternately, use of the shift-carbonation process could be limited to gasification processes that do not produce significant quantities of \( N_2 \) and \( CH_4 \).

The second major problem is the high energy requirement of the endothermic sorbent regeneration reaction. Although the energy released by the exothermic shift and carbonation reactions is approximately equal to the energy required for regeneration, the higher regeneration temperature makes energy exchange impossible. Combustion of approximately 50% of the raw coal gas, which is required to satisfy the regeneration energy requirement, is directly responsible for the reduced \( H_2 \) production rate. Alternate energy sources, such as waste gas from nearby processing units, or heat integration with the gasification process are required.

Finally, the pressure difference between the shift-carbonation reactor (340 psia) and the regeneration reactor (19 psia) creates a significant problem. Dual recirculating fluidized-bed reactors, similar to those used in the fluidized-bed catalytic cracking process, are an obvious choice for transporting the large quantities of sorbent between the two reactors. Coupled fluidized beds, however, must operate at approximately equal pressures. Significant reduction in the shift-carbonation reactor pressure is unacceptable since the effectiveness of the carbonation reaction decreases at low pressure. Increasing the regeneration reactor pressure is undesirable because higher regeneration temperature, which would likely accelerate sorbent deterioration, would be required. Innovative thinking will be required to solve the solid transport problem or to identify alternate reactor configurations which can function under the pressure difference and still achieve the desired rapid solid transfer.

**Alternate Applications**

There are other applications where the combined reaction-separation process may be applied. One possibility involves carrying out the steam-methane reforming reaction in the presence
of calcined dolomite. This process, which has been suggested previously by Brun-Tsekhovsi et al. (1988), would enable the reforming, shift, and primary CO$_2$ separation steps to be accomplished in a single vessel. On the basis of equilibrium calculations, a reactor product containing 95% H$_2$ (dry basis) could be produced at 1000 K and 15 atm with a feed consisting of 4 mols of steam per mol of CH$_4$. The equilibrium yield of H$_2$ would be 3.5 mols per mol of CH$_4$ fed, or 88% of the stoichiometric maximum. Impurities would consist of about 3% unreacted CH$_4$ and 1% of both CO and CO$_2$ not removed in the shift-carbonation step. Further purification to produce essentially pure H$_2$ could be accomplished using PSA. In this concept, the endothermic reforming and exothermic shift and carbonation reactions would occur at the same temperature to produce an approximate energy balance. Energy from the combustion of CH$_4$ would be required for sorbent regeneration. In the current reforming process, combustion of large quantities of CH$_4$ is required to drive the endothermic reforming reaction.

Traditional steam-methane reforming at the same steam to CH$_4$ ratio, temperature, and pressure would, at equilibrium, produce a reformer product containing only 71% H$_2$ (dry) basis, 61% of the theoretical yield. Subsequent processing including multistage catalytic shift reactors and CO$_2$ separators would be required to increase the H$_2$ content and yield to the values obtained in the combined reforming-shift-carbonation process.

In another possible application, the combined shift-carbonation process might be coupled with a coal gas-fired molten carbonate fuel cell operating at approximately 1 atm and 650°C. There are two possible locations for the shift-carbonation step. If the coal gas is fed to the shift-carbonation reactor before entering the anode, the H$_2$ content would be increased and the CO$_2$ content decreased. Both would improve anode operation. Sorbent regeneration would produce the CO$_2$ required as cathode feed, which is now produced by combustion of a portion of the coal gas.

Alternately the sorbent process could be located at the anode exit. Only the carbonation reaction would be important since the shift reaction would go essentially to completion within the anode. Selective removal of CO$_2$ from the anode exhaust would permit unreacted H$_2$ to be recycled to the anode so that 100% of the fuel would be used to generate electricity. Again, CO$_2$ obtained during the sorbent regeneration step would be fed to the cathode. The molten carbonate fuel cell application has the advantage of operating near atmospheric pressure, where the dual recirculating fluidized-bed reactor concept could be applied.

**FUTURE WORK**

This contract ended in December 1994. The final report has been submitted, and it should soon be published.

**REFERENCES**


Han, C. and D.P. Harrison, 1994, Simultaneous Shift Reaction and Carbon Dioxide Separation for the Direct Production of Hydrogen, Chemical Engineering Science, 49, 5875.

Harrison, D.P., et al., 1993, A Calcium Oxide Sorbent Process for Bulk Separation of Carbon Dioxide, IV., Proceedings of the Symposium on Coal Fired Power Systems ’93,


