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**Investigation of Processes to Treat Zinc-
Ferrite Regeneration Offgases in High
Temperature Desulfurization of Coal Gases**

Final Report

L.R. Woodland

August 1988

Work Performed Under Contract No.: DE-AC21-86MC23086

**For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia**

**By
Arthur D. Little, Inc.
Cambridge, Massachusetts**

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The contributions of Mr. Thomas Dorchak, U.S. Department of Energy, are gratefully acknowledged. His efforts in program formulation and critical review of results constituted a vital part of the overall effort.

ABSTRACT

A bench-scale test program has been performed to evaluate the effect of critical operating variables on sulfur dioxide removal efficiency for two candidate "throwaway" desulfurization processes for treating zinc-ferrite reactor regeneration offgas. These two "throwaway" processes, dual-alkali wet scrubbing and spray dryer desulfurization, generate a gypsum waste sludge environmentally acceptable for disposal. They have seen extensive commercial application in flue gas desulfurization, but have not been evaluated for such high levels of removal efficiency and such high inlet sulfur dioxide concentrations as would be required for this application.

In this bench-scale test program, it was demonstrated that the dual-alkali desulfurization system is capable of reducing inlet sulfur dioxide concentrations of up to 120,000 ppm to below 100 ppm, while the spray dryer desulfurization system is limited to treating inlet sulfur dioxide concentrations of up to 15,000 ppm to less than 100 ppm. The spray dryer desulfurization system has the additional disadvantage of not allowing heat recovery for gas stream recycle to the zinc-ferrite reactor, since all the heat in the entering gas stream is required for water evaporation.

The estimated capital cost for the two "throwaway" desulfurization systems for servicing a 100 MWe coal gasifier are approximately \$8.6 million for the dual-alkali system and \$15 million for the spray dryer system. This compares with about \$27 million for the Wellman-Lord sulfur recovery system.

The spray dryer system has a much higher capital cost than the dual-alkali system because the spray dryer system must also include a large investment for particulate collection

(for collecting spent alkali particulate from the spray dryer) and a solids recycle system necessary for limiting excess alkali requirements. The Wellman-Lord sulfur recovery process has a high capital cost because it is a very complex multistep chemical process.

The net operating costs for these three systems are estimated at about 10 mills/kWh for the dual-alkali system, 16 mills/kWh for the spray dryer system, and 21 mills/kWh for the Wellman-Lord system.

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1.0 EXECUTIVE SUMMARY

Morgantown Energy Technology Center (METC) has been developing processes for high-temperature desulfurization of coal gases. METC has done considerable work on the zinc-ferrite high-temperature gas desulfurization process. This process offers several advantages including extremely high sulfur capture and operating temperatures that are consistent with all contemplated end uses. The high-temperature capture of sulfur from coal gas by any reusable sorbent requires regeneration. METC has been considering several regeneration options. Regeneration of zinc-ferrite or other metal oxide sorbents is critical to the successful implementation of a technique for the high-temperature desulfurization of coal gases.

The regeneration offgas that results from the passage of oxidation air through spent desulfurization sorbent beds contains primarily nitrogen (88 to 97.5%) and sulfur dioxide (2.5 to 12%). The regeneration offgas may also contain small amounts of elemental sulfur, hydrogen sulfide, ammonia, hydrogen chloride, sulfur trioxide, and trace volatile metals. The gas temperature can reach as high as 1500°, but at the end of the regeneration step, the temperature in the reactor may fall to about 1000°F. Pressure is usually in the order of 300 psig. The composition of the regeneration offgas may also vary during the regeneration cycle.

The enthalpy content in the nitrogen and/or steam in the offgas in the regeneration process is considerable; therefore, there is incentive to recycle the regeneration gas after appropriate sulfur control. However, for the gasification system to operate properly, the sulfur dioxide in the regeneration offgas would have to be removed to

sufficiently low levels, say 100 ppm, to avoid sulfation of the hot gas sorbent and preclude other operational problems.

Of the two processes tested in our bench-scale testing program, the dual-alkali (wet scrubbing) desulfurization system operated in conjunction with heat recovery would meet the above stated requirements of energy conservation and high efficiency. A six-stage (tray) dual-alkali absorber can reduce inlet SO_2 concentrations of up to 120,000 ppm to below 100 ppm operating at near atmospheric pressure. (Operating at higher pressure - 15 psig or greater - can be used to reduce the number of trays necessary or further increase SO_2 removal efficiency.)

The water vapor introduced in the dual-alkali absorber would be an acceptable diluent (in place of some of the nitrogen from regeneration air) in the recycle stream to the zinc-ferrite reactor. The water vapor or nitrogen acts to control regeneration temperature in the reactor. If necessary, the volume of water vapor introduced in the dual-alkali absorber could be limited by pressure operation.

Unlike the dual-alkali desulfurization system, the spray dryer desulfurization system would be limited to treating zinc-ferrite offgas of up to only 15,000 ppm SO_2 . All the available enthalpy in the zinc-ferrite offgas stream would be required for water evaporation in the spray dryer at a 15,000 ppm SO_2 inlet concentration, so there would be little incentive for recycle of this gas stream to the zinc-ferrite reactor. The estimated capital cost for the two "throwaway" desulfurization systems for servicing a 100 MWe coal gasifier are approximately \$8.6 million for the dual-alkali system and \$15 million for the spray dryer system. This compares with about \$27 million for the Wellman-Lord sulfur recovery system.

The estimated capital cost for the two "throwaway" desulfurization systems for servicing a 100 MWe coal gasifier are approximately \$8.6 million for the dual-alkali system and \$15 million for the spray dryer system. This compares with about \$27 million for the Wellman-Lord sulfur recovery system. The capital investment for the spray dryer desulfurization system is considerably higher than for the dual-alkali absorption system primarily because of the requirements for particulate collection (of the spent alkali) and recycle of alkali solids to increase alkali utilization. The Wellman-Lord sulfur recovery process has a higher capital investment because it is a complex, multi-step chemical process.

The net operating costs for these three systems are estimated at about 10 mills/kWh for the dual-alkali system, 16 mills/kWh for the spray dryer system, and 21 mills/kWh for the Wellman-Lord system.

Whether or not this "throwaway" desulfurization process would be less expensive than the new concept for recycle of zinc-ferrite regeneration offgas back to the fluid bed coal gasifier, however, remains to be determined.

2.0 BACKGROUND AND INTRODUCTION

Morgantown Energy Technology Center (METC) has been developing processes for high-temperature desulfurization of coal gases. METC has done considerable work on the zinc-ferrite high-temperature gas desulfurization process. This process offers several advantages including extremely high sulfur capture and operating temperatures that are consistent with all contemplated end uses. The high-temperature capture of sulfur from coal gas by any reusable sorbent requires regeneration. METC has been considering several regeneration options. Regeneration of zinc-ferrite or other metal oxide sorbents is critical to the successful implementation of a technique for the high-temperature desulfurization of coal gases.

The regeneration offgas that results from the passage of oxidation air through spent desulfurization sorbent beds contains primarily nitrogen (88 to 97.5%) and sulfur dioxide (2.5 to 12%). The regeneration offgas may also contain small amounts of elemental sulfur, hydrogen sulfide, ammonia, hydrogen chloride, sulfur trioxide, and trace volatile metals. The gas temperature can reach as high as 1500°, but at the end of the regeneration step, the temperature in the reactor may fall to about 1000°F. Pressure is usually in the order of 300 psig. The composition of the regeneration offgas may also vary during the regeneration cycle.

The enthalpy content in the nitrogen and/or steam in the offgas in the regeneration process is considerable; therefore, there is incentive to recycle the regeneration gas after appropriate sulfur control. However, for the gasification system to operate properly, the sulfur dioxide in the regeneration offgas would have to be removed to

sufficiently low levels, say 100 ppm, to avoid sulfation of the hot gas sorbent and preclude other operational problems.

The DOE has sponsored work to develop a process that will be able to recover elemental sulfur from the zinc-ferrite regeneration offgas stream. The economics of this process, however, do not appear favorable. The DOE would like to develop a cost-effective, non-recovery process technology for this application similar to the conventional commercial flue gas desulfurization (FGD) processes.

Implementation of conventional sulfur dioxide FGD technology has been growing very rapidly for coal fired power plants in the United States. By late 1985, more than 50,000 megawatts of capacity had FGD systems in operation in the United States. Two categories of FGD technologies are generally in use:

- (1) Non-recovery systems, which produce a solid or liquid waste with little market value at present; and
- (2) Recovery systems, which produce elemental sulfur or sulfuric acid as a by-product for sale.

At present, the overwhelming majority of FGD systems for controlling emissions from industrial and utility boilers use the non-recovery or "throwaway" processes.

There are a number of non-recovery FGD processes being used commercially for treatment of flue gas from pulverized coal-fired boilers that could be used for removal of sulfur dioxide from the desulfurization reactor regeneration offgas. The more common of these commercial FGD systems are as follows:

Wet Scrubbing

Lime

Limestone

Alkaline Fly Ash

Soluble Alkali (Sodium Base)

Dual-Alkali (Sodium/Lime)

"Dry" (Spray Dryer)

Absorption

Sodium (Soda Ash or
Nacholite)

Lime

Limestone

The first three of these wet scrubbing processes utilize slurries of lime, limestone, or fly ash to remove sulfur from flue gases. They produce slurries containing 5-20% solids which are further dewatered and processed prior to placement in landfills (or they may be ponded). Landfilling of the FGD wastes (instead of ponding) is likely to prove the more popular waste disposal method in the future. The soluble alkali FGD processes (usually sodium base) produce soluble sulfite or sulfate salts that pose a disposal problem except in arid areas where the scrubbing solutions can be evaporated in ponds for removal and disposal of the salts. The double-alkali process is a second-generation technology that utilizes solutions of sodium salts for SO₂ removal. These solutions are then regenerated, using lime or limestone, to produce an insoluble, calcium-based waste solid that is discharged as a filter cake.

Three approaches for dry FGD control have been studied:

- injection of absorbent slurries in a spray dryer and collection in a baghouse;
- injection of solid sorbent in the flue gas stream and the downstream collection; and

- direct injection of absorbent materials in the combustion zone of the furnace.

The most reliable commercial dry FGD sulfur control technology today features the injection of absorbent slurries in a spray dryer and the collection of waste in a baghouse or electrostatic precipitator. Lime and soda ash have been the typical sorbents employed.

The older wet scrubbing systems are usually higher in efficiency (and cost) than are the newer spray dryer absorber systems in which a water slurry of alkali is injected into the flue gas and the dried sorbent product collected in a baghouse or ESP.

A simplified block flow diagram showing how the hot gas desulfurization system and the proposed FGD process tail gas cleanup system could be integrated is presented below:

During the hot coal gas desulfurization cycle, the zinc-ferrite mixed oxide would react with hydrogen sulfide in the gas stream to form mixed sulfides and water vapor. When

the zinc-ferrite bed in the reactor became saturated with sulfur, the bed would be taken off-line and regenerated using makeup air (for oxygen) and recycled treated offgas, consisting primarily of nitrogen and water vapor, as diluent.

Two zinc-ferrite reactors would be used to allow one to be on-line for desulfurization of hot coal gas, while the second was being regenerated in conjunction with operation of the desulfurization system.

Heat recovery would be utilized prior to the offgas desulfurization system, where possible, to conserve energy. The end product of the treatment system should be an environmentally acceptable waste for disposal, such as insoluble calcium salts.

3.0 PURPOSE

The main objective of this bench-scale test program was to evaluate the critical, unproven elements of two commercial flue gas desulfurization (FGD) processes, dual-alkali and spray drying, as they might be applied to treatment of zinc-ferrite regeneration offgas. The most critical unproven element for both of these FGD processes was whether sufficient sulfur dioxide removal efficiency could be achieved and what operating conditions would be required. A second objective of this program was to use the data obtained in the bench-scale test program to generate a conceptual design for a full-scale (100 MWe) desulfurization system for each of these two processes. The final objective was to prepare capital and operating cost estimates for these two processes and compare them with a sulfur recovery desulfurization process.

Although these two processes have had extensive commercial application for flue gas desulfurization with coal-fired utility boilers, it has been for much lower sulfur dioxide concentrations (3-4000 ppm sulfur dioxide) and at lower removal efficiencies (70-95%). (To reduce a 120,000 ppm inlet sulfur dioxide concentration to below 100 ppm, outlet concentration requires a removal efficiency of over 99.9%.)

The operating conditions investigated in our bench-scale (10-20 scfm) system for the dual-alkali desulfurization system included the following:

- Alkali/SO₂ ratio
- Alkali pH
- Operating pressure
- Gas velocity

Inlet sulfur dioxide concentration
Number of operating trays (transfer stages)

The operating conditions investigated for the spray dryer system included alkali-to-sulfur dioxide ratio and approach to saturation temperature.

The data generated in testing these two bench-scale desulfurization systems was used to design a conceptual full-scale system. Capital and operating cost estimates were then prepared for these two desulfurization processes and one sulfur recovery process (Wellman-Lord).

4.0 DESIGN OF BENCH-SCALE TESTING SYSTEMS

Both bench-scale desulfurization systems were supplied with synthetic zinc-ferrite regeneration offgas from the gas supply system as shown in Figure 4-1.

Hot gas was supplied at 10-20 scfm to the dual-alkali absorber (Figure 4-2) or spray dryer (Figure 4-3). The source of the nitrogen gas was a liquid nitrogen DeWar, a vaporizer and a pressure regulator. The source of sulfur dioxide was a two-cylinder gas manifold and pressure regulator. The flows of nitrogen and sulfur dioxide were measured by rotameters at 10-30 psig. The nitrogen was first passed through a preheater to bring it up to 400-500°F. The sulfur dioxide was introduced into the nitrogen gas stream between the nitrogen preheater and the main (high temperature) multipass tubular heater, where the two gas streams were mixed and heated to 400-1000°F.

The dual-alkali absorber (Figure 4-2) consisted of two sections. The bottom section was a 2 ft high x 6 inch diameter stainless steel quench section equipped with a quench spray nozzle for cooling the inlet gas from about 400°F to about 130°F. The upper 8 ft high x 6 inch (inside) diameter Plexiglass® section contained six sieve trays, each tray having a total of 704 holes one-sixteenth inch in diameter. Each tray also had two 1-inch diameter "downcomer" tubes for channeling the absorber liquid from each tray to the one below and one-half inch inlet and outlet overflow weirs to maintain the liquid/gas seal and liquid depth on each tray.

A heating tape traced and insulated manifold was connected to sample valves located on the inlet gas pipe, above the quench spray, above and below each tray, and in the absorber outlet duct to allow measurement of sulfur dioxide concentrations

FIGURE 4-1

SCHEMATIC DIAGRAM OF SYNTHETIC OFFGAS SUPPLY SYSTEM

FIGURE 4-3

SCHEMATIC DIAGRAM OF SPRAY DRYING SULFUR REMOVAL SYSTEM

throughout the absorber system. Sodium sulfite feed solution (from the 250 gal feed solution tank) was heated to 130°F in a steam-jacketed heat exchanger prior to being fed to the quench spray or trays to simulate the conditions that would be present in a commercial system where the solution would be returned hot (about 130°F) from the thickener. Sodium sulfite solution could be introduced to either the fourth or sixth tray as desired. A float control switch installed in the bottom of the quench tank operated a solenoid absorber liquid discharge valve allowing operation of the absorber under pressure (up to 30 psig). Spent absorber liquid was collected in one of two 250-gallon spent liquor tanks.

The spray dryer (as shown in Figure 4-3) was a 31-inch pilot (portable) unit, leased from Niro Atomizer Co. of Columbia, MD, which was equipped with a rotary turbine-driven atomizer, a cyclone solids collector and induced draft fan. (A heater is also normally supplied with this portable spray dryer, but for our bench-scale testing this was replaced by the synthetic offgas supply system.) A 5-gallon glass carboy was connected to the solids collection cyclone in place of the 1-quart solids collection jar supplied with the spray dryer, to provide much greater solids capacity and allow a longer operating period for the spray dryer system to reach equilibrium.

A specially modified DuPont 400 Photometric (Ultraviolet) SO₂ Analyzer was used for measuring SO₂ concentration from any of the sampling points during the bench-scale testing of the dual-alkali absorber and spray dryer desulfurization systems. (An ultraviolet analyzer was chosen rather than an infrared analyzer because water vapor will interfere with the infrared measurement of SO₂.) The modified DuPont 400 Analyzer was a dual range unit measuring 0-1000 and 0-120,000 ppm SO₂. (The unmodified standard unit has a single range.)

It also had a rapid response time (about 10 seconds) which enabled a complete set of gas samples to be taken at the scrubber inlet and outlet and above and below each tray within a 10-minute time span during the testing of the dual-alkali absorber.

5.0 OPERATION OF THE BENCH-SCALE DESULFURIZATION SYSTEM

5.1 Synthetic Offgas Supply System

Prior to starting gas flow through the synthetic offgas supply system, the nitrogen gas preheater and combined gas (tubular) electric heaters were turned on and allowed to come up to operating temperature. Nitrogen gas flow was then started and water flow introduced to the (dual-alkali scrubber or spray dryer) FGD system. When the operating (nitrogen) gas temperature was established, injection of SO_2 was begun and its concentration in the synthetic offgas confirmed by the on-line SO_2 analyzer. (At the same time the SO_2 flow was started, alkali introduction to the FGD unit was also begun.) When the appropriate gas analyses had been made across the FGD unit at one inlet gas concentration (e.g., 3% SO_2), the SO_2 injection rate was changed to the next inlet concentration (e.g., 6%).

5.2 SO_2 Analyzer

Before each run, the SO_2 analyzer was calibrated over both measuring ranges, 0-1000 and 0-120,000 ppm, using standardized calibration gas mixtures of SO_2 in nitrogen supplied by Matheson Gas Co. of Gloucester, MA. The 0-1000 ppm range was calibrated using 65 and 675 ppm SO_2 cal. gas and the 0-120,000 ppm range calibrated using 9,700 and 68,000 ppm SO_2 cal. gas.

5.3 Dual-Alkali Absorber

To conserve alkali solution, water was initially fed to the absorber trays and quench spray while the nitrogen gas stream came up to temperature. When the nitrogen gas stream reached temperature and the introduction of SO_2 was begun, feed of

the sodium sulfite solution in place of the water was also started.

Usually about 15-20 minutes after sulfur dioxide introduction was started, the system reached equilibrium and gas sampling was started. Complete collection of gas samples normally required about 10-15 minutes. This included sampling the absorber inlet and outlet, after the quench and above and below each tray.

5.4 Spray Dryer

As in the case of the dual-alkali absorber, the spray dryer was started up using water (supplied by peristaltic pump) at 40-80 cc/min until the spray dryer came up to temperature (with inlet nitrogen gas at about 1000°C). When the spray dryer reached operating temperature (inlet about 1000°F, outlet about 200°F), sulfur dioxide injection into the nitrogen gas stream and alkali slurry (or solution) feed to the spray dryer atomizer were started simultaneously. While the inlet SO₂ concentration was held constant at about 1%, the alkali feed rate was varied to change alkali-to-SO₂ ratio and the approach to saturation in the outlet gas.

6.0 BENCH-SCALE TESTING DATA ANALYSIS

The basic test data collected during the bench-scale testing of the dual-alkali absorber are summarized in Tables 6-1 through 6-5. The test data collected during the operation of the bench-scale spray dryer desulfurization system are summarized in Table 6-6 and Figures 6-1 to 6-3.

6.1 Initial Analysis of Dual-Alkali Absorber Testing

Scoping test runs were made initially to set three of the basic operating parameters: quench spray L/G ratio, absorber tray L/G ratio, and gas velocity.

The quench spray, in addition to lowering the inlet gas temperature (from about 400°F to about 140°F), also contributes to sulfur dioxide removal efficiency. Table 6-1 summarizes the effects of quench spray and tray L/G on SO₂ removal efficiency. As can be seen in Table 6-1, the combined SO₂ removal efficiencies of the quench spray and first tray (which is located just over the quench spray) increased with the increase in quench spray L/G from 8 to 24. There appeared to be much less increase in combined SO₂ removal efficiency with the increase in quench spray L/G to 32. A quench spray L/G of 24 (rather than the L/G = 16 originally proposed in the test plan) was, therefore, used in all subsequent testing.

The influence of liquor flow rate (L/G) to the trays is presented in Table 6-2. For 3% sulfur dioxide input to the absorber and no SO₂ removal by the quench spray, the L/G to the trays necessary to supply the theoretical weight of alkali needed for 100% SO₂ removal is 40. With 40% SO₂ removal by the quench spray, however, the theoretical L/G to the trays necessary for 100% SO₂ removal is reduced to 24.

TABLE 6-1

SUMMARY OF DUAL-ALKALI ABSORBER QUENCH FLOW RATE (L/G) RUNS DAQ1-DAQ4

Basis: 10 scfm, 6% SO₂, Tray L/G* = 77, 0.5N Na₂ SO₃ liquor

Run	Quench L/G*	Overall Efficiency Tray + Quench	Quench Efficiency	First Tray Efficiency	Quenched Gas Temp. (°F)
DAQ1	8	63.6%	13.8%	57.8%	144
DAQ2	16	73.5	28.7	62.8	148
DAQ3	24	79.5	39.5	66.1	143
DAQ4	32	82.1	47.8	65.8	130

TABLE 6-2

SUMMARY OF DUAL-ALKALI ABSORBER TRAY FLOW RATE (L/G) RUNS DAT1-DAT2

Basis: 10 scfm, Quench L/G* = 24, 3% SO₂, 0.5N Na₂ SO₃ liquor

Run	Tray		SO ₂ Removal Efficiency		Average of Tray 1 & 2
	Flow (gpm)	L/G*	Quench + Tray 1	Tray 2	
DAT1	0.77	77	82.0%	60.0%	71%
DAT2	0.58	58	82.0	69.9	76
DAT3	0.48	48	77.3	70.0	74
DAT4	0.38	38	72.6	64.8	69

*L/G = liquor to gas ratio, gpm/1000 scfm

TABLE 6-3

SUMMARY OF DUAL-ALKALI ABSORBER TRAY GAS VELOCITY TEST RUNS (DA19-21)6 Trays, 0.5N Na₂SO₃ liquor, Tray L/G = 77, Quench L/G = 24, 6% SO₂

Run	Tray Gas Velocity		Column Pressure Drop (in.W.C.)	Inlet SO ₂ Conc. (% SO ₂)	Liquor Flow gpm	Tray Efficiencies (Percent Removal)						Outlet ppm SO ₂
	scfm	ft/sec				Q + 1*	2	3	4	5	6	
DA21	10	11.1	6.2	6.2	0.38	57.7	57.2	70.1	72.1	73.4	74.0	65
DA20	15	16.6	7.9	6.5	0.57	46.2	46.9	55.8	68.5	67.0	65.9	300
DA19	20	22.2	9.4	6.0	0.76	38.6	36.7	47.2	59.5	68.3	64.8	500

*Quench + 1st tray combined

TABLE 6-4

SUMMARY OF DUAL ALKALI ABSORBER ATMOSPHERIC PRESSURE RUNS DA1-DA12
(All runs 10 scfm inlet gas)

<u>Run</u>	<u>Trays</u>	<u>% SO₂</u>	<u>Tray Liquor</u>		<u>Inlet pH</u>	<u>Tray Efficiencies (Percent Removal)</u>							<u>Outlet ppm SO₂</u>
			<u>L/G*</u>	<u>Conc.</u>		<u>Q+1**</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>6</u>	<u>Overall</u>	
DA1	6	3.3	38	0.5N	9.4	76.1%	65.6%	71.5%	72.0%	53.5%	70.0%	99.91%	30
DA2	6	6.3	77	0.5N	9.4	63.4	64.2	68.2	69.7	77.5	61.1	99.89	70
DA3	6	12.0	169	0.5N	9.4	56.5	60.7	75.1	81.3	84.2	56.7	99.95	65
DA4	6	3.1	38	0.5N	12.0	78.7	67.2	68.3	77.9	66.7	50.0	99.92	25
DA5	6	6.2	77	0.5N	12.0	72.8	65.9	70.7	67.9	66.7	83.3	99.95	30
DA6	6	12.2	169	0.5N	12.0	62.7	65.8	76.8	77.1	66.3	85.7	99.97	40
DA7	6	3.1	38	1.0N	9.5	83.5	67.8	72.9	68.9	67.1	56.5	99.90	20
DA8	6	6.9	77	1.0N	9.5	82.8	72.7	69.3	74.7	72.0	71.4	99.97	20
DA8A	6	7.1	48	1.0N	9.5	79.3	67.0	68.6	68.6	75.0	75.0	99.96	30
DA9	6	14.2	120	1.0N	9.5	68.3	71.5	69.5	65.0	70.0	84.0	99.95	65
DA10	4	3.2	38	0.5N	9.4	79.8	64.8	73.4	75.0	--	--	99.50	150
DA11	4	6.4	77	0.5N	9.4	69.2	63.9	67.1	72.4	--	--	99.00	640
DA12	4	11.9	169	0.5N	9.4	59.9	65.1	73.9	80.9	--	--	99.30	820

*Gpm scrubbing liquor per 1000 scfm inlet gas. Quench L/G = 24.

**Quench + 1st tray combined.

TABLE 6-5

COMPARISON OF HIGH AND LOW PRESSURE ABSORBER DUAL-ALKALI RUNS
(DA1-3, 13-15, 16-18, 22-24)*

Run	Inlet Gas			Tray L/G	Tray Gas Velocity (ft/sec)	Tray Efficiencies (Percent Removal)						Overall	Outlet ppm SO ₂
	% SO ₂	Flow (scfm)	Pressure (psia)			Q + 1	2	3	4	5	6		
Base (Low Pressure) Runs - 6 Trays - Variable Inlet SO ₂ Concentration													
DA1	3.3	10	15	38	11.1	76.1	65.6	71.5	72.0	53.5	70.0	99.91	30
DA2	6.3	10	15	77	11.1	63.4	64.2	68.2	69.7	77.5	61.1	99.89	70
DA3	12.0	10	15	169	11.1	56.5	60.7	75.1	81.3	84.2	56.7	99.95	65
High Pressure Runs - 6 Trays - Variable Inlet SO ₂ Concentration													
DA13	2.8	10	30	38	5.6	80.9	82.0	70.8	90.4	77.8	>90	100.00	<1
DA14	6.2	10	30	77	5.6	74.8	83.0	75.0	92.4	70.0	>93	100.00	<1
DA15	11.5	10	30	169	5.6	53.9	69.0	89.5	95.4	87.5	>90	100.00	<1
High Pressure Runs - 4 Trays - Variable Inlet SO ₂ Concentration													
DA16	3.2	10	30	38	5.6	54.5	70.0	89.9	94.3	--	--	99.72	90
DA17	6.6	10	30	77	5.6	68.3	81.2	82.0	87.1	--	--	99.86	90
DA18	11.5	10	30	169	5.6	80.3	79.0	71.6	76.0	--	--	99.92	90
High Pressure Runs - 6 Trays - Variable Gas Velocity													
DA22	3.2	20	30	38**	11.1	64.3	54.8	61.8	72.8	69.8	68.8	99.84	50
DA23	3.2	10	30	38	5.6	68.4	84.0	71.3	67.4	73.3	75.0	99.97	10
DA24	2.8	10	44	43	3.7	64.6	86.7	79.5	74.1	80.0	92.8	100.0	1

Notes: *0.5N Na₂SO₃ scrubbing solution, Inlet Liquor pH = 9.4, Quench L/G = 30 unless otherwise noted

**Quench L/G = 55

TABLE 6-6

SUMMARY OF SPRAY DRYER ABSORPTION TEST RUNS

Test Run Designation	Type Alkali	Alkali/SO ₂ Ratio	SO ₂ Concentration (Dry Gas Basis)		Gas Inlet Temp. (°F)	Gas Outlet Temp. (°F)	Approach to Sat.Temp. (ΔT_{as})	% Removal Efficiency (Dry Gas Basis)
			Inlet (ppm)	Outlet (ppm)				
LS-1	Hydrated Lime	1.02	10,200	5345	980	273	141°F	47.6
		1.15	10,000	4770	990	264	132	52.3
		1.30	9,900	4640	1020	248	116	53.1
		1.38	9,900	4400	1010	255	123	55.6
		1.38	9,900	4050	1010	239	107	59.1
SA-2	Soda Ash	1.13	10,060	880	1000	222	85	91.3
SA-4		1.18	9,600	1010	1010	241	106	89.5
SA-4		1.05	9,600	1085	1000	253	121	88.7
SA-3		0.957	11,200	1390	946	227	95	87.6
SA-3	Slaked Quicklime	0.85	11,200	1950	923	257	127	82.6
LS-2		1.91	11,030	3086	1050	208	65	70.0
		2.23	11,030	2680	1040	198	51	75.7
		2.55	11,030	990	1060	180	29	93.0
		2.88	11,030	84	1080	165	5	99.2

Figure 6-1 Effect of Alkali/SO₂ Ratio on Spray Dryer SO₂ Removal Efficiency Using 30%(by wt.) Hydrated Lime Slurry

Figure 6-2 Effect of Alkali/SO₂ Ratio on Spray Dryer SO₂ Removal Efficiency Using 35%(by wt.) Soda Ash Solution

Figure 6-3 Combined Effect of Alkali/SO₂ Ratio and Approach to Saturation on Spray Dryer SO₂ Removal Efficiency

An L/G of 38 to the trays, therefore, already supplies about 60% excess alkali. (Excess active alkali to the absorber in a dual-alkali system is not lost, as in the case of "once-through" scrubbing system, but merely increases the recycle flow rate in the system.) Increasing the L/G beyond 38 (more than 60% excess alkali) further increases SO₂ removal efficiency, but only marginally. It also appears that when the L/G is increased from 58 to 77, SO₂ removal efficiency either stabilizes or is actually decreased because of the high hydraulic loading and resultant approach to column flooding. An L/G ratio of 38 to the trays was, therefore, considered sufficient for a 3% inlet SO₂ concentration. To maintain this same alkali-to-SO₂ ratio as the inlet SO₂ concentration was increased, the L/G was increased to about 77 for 6% SO₂ and 169 for 12% SO₂.

The gas sampling points above and below each of the trays were fixed. There was some variation in the gas flow pattern through each of the trays, however, and this sometimes resulted in fluctuating SO₂ concentration measurements, particularly over the first or second tray. By the time the gas flow reached the top tray (#6), there was less variation in the gas flow pattern and in the SO₂ readings. For this reason, the readings for the quench (Q) and trays 1 and 2 were averaged in Table 6-1 for Runs DAT1-4, and this appeared to give more consistent results.

In Table 6-3 the effect of gas velocity through the trays (at atmospheric pressure and constant L/G) is examined. As the gas velocity is increased from 11 ft/sec (for 10 scfm) to 22 ft/sec (for 20 scfm), there is a marked reduction in overall and tray by tray SO₂ removal efficiency. Because the 10 scfm gas rate (11 ft/sec gas velocity through the trays) gave a markedly better SO₂ removal efficiency and also met overall SO₂ removal efficiency requirements of less than 100 ppm

outlet concentration, the 10 scfm gas rate was used for all subsequent dual-alkali test runs. (The original test plan had proposed a gas flow rate of 20 scfm based on our commercial system experience.)

The pressure drop for these three runs (DA19-21) ranged from 6.2 to 9.4 inches water column (W.C.). While this pressure drop would represent a significant cost (capital and operating cost) in a conventional FGD system which must treat a very large volume of flue gas, it would be much less significant in this application where the gas volume would be 10-40 times less and the pressure drop is only a small fraction of the pressure leaving the zinc-ferrite reactor.

The dual-alkali test runs summarized in Table 6-4 were run under somewhat different test conditions than those originally outlined in the Test Plan as the result of the scoping tests (which were presented in Tables 6-1 and 6-3). These changes in operating conditions were as follows:

- Total gas flow rate was changed from 20 to 10 scfm;
- Quench liquor flow rate was changed from $L/G = 16$ to $L/G = 24$;
- Tray flow rates were changed from $L/G = 30, 60, 120$ to 38, 77, 169, respectively; and
- The pH of 0.5N commercial Na_2SO_3 was 9.4.

The test variables examined in the runs reported in Table 6-4 were as follows:

- Three inlet SO_2 concentrations: 3, 6 and 12% sulfur dioxide;
- Inlet liquor pH of 9.4-9.5 and 12.0;
- Inlet liquor alkali concentrations of 0.5 and 1.0 N active sodium; and

- Four or six operating absorber trays.

The tests were all run at a gas flow rate of 10 scfm and a quench liquor flow rate of 0.24 gpm (an L/G of 24). The absorber liquor was preheated to 130°F before being fed to the absorber to simulate hot liquor return from a full-scale dual-alkali regeneration system.

The liquid to gas ratio, L/G, (gpm/1000 acfm) to the trays was changed for each inlet sulfur dioxide concentration based on a 50-60% excess feed alkali over the stoichiometric quantity required for complete sulfur dioxide removal.

Tests DA1-DA3 in Table 6-4 show, that with six trays, an overall efficiency of 99.90-99.95% and an outlet concentration of less than 100 ppm can be achieved for a 3-12% sulfur dioxide input using 0.5N Na_2SO_3 and a quench spray rate of LG=24. When the pH of the 0.5N Na_2SO_3 is increased to 12.0 by replacing 2% of the Na_2SO_3 in the scrubbing liquor with NaOH, the overall scrubbing efficiency is increased to 99.92-99.97% and the average outlet concentration is reduced from about 60 ppm to about 30 ppm.

When the available alkali feed rate is doubled to give about 200% excess (by doubling the feed liquor concentration of active Na^+ from 0.5N to 1.0 N), higher scrubbing efficiencies (99.90-99.97%) are again achieved along with a lower average outlet concentration of about 30 ppm. This higher alkali concentration gives roughly the same effect as operating with a high pH (12.0), 0.5N active Na^+ scrubbing liquor.

Operation with only four trays at atmospheric pressure does not produce an overall sulfur dioxide removal efficiency sufficient to obtain the necessary 100 ppm or less in the outlet gas stream.

Table 6-5 presents a summary of the data collected during the pressure operation of the absorber in comparison to (base) Runs DA1-3 which were made at atmospheric pressure. High pressure test runs were made at 15 psig as called for in the test plan, with one at an even higher pressure of 29 psig as well. The data presented for Runs DA13-18 in Table 6-5 shows a marked improvement in SO₂ removal efficiency for pressure operation at 15 psig at all three SO₂ inlet gas concentrations. It also indicates that the desired outlet SO₂ concentration of 100 ppm or less could be achieved with only four trays rather than the six necessary at atmospheric pressure operation.

Operation at 29 psig (DA24) may have increased the SO₂ removal efficiency beyond that achieved with 15 psig operation (DA23), although this is not clear from the data. If DA24 is compared with DA13 instead of DA23 (both of the latter runs under the same operating conditions), there does not appear to be any improvement. We cannot account for the difference in results for Runs DA13 and 23. Unfortunately, the discrepancy between Runs DA13 and DA23 was not discovered until after all testing with the dual-alkali systems was completed and disassembly of the bench-scale unit started; otherwise another run would have been made under these conditions. Nevertheless, we believe that results within subsets of data (e.g., Test Runs DA13-15 and DA22-24) are more reliable since such variables as SO₂ analyzer accuracy and scrubber liquor composition are more likely to be consistent over these shorter periods of time.

Two possible reasons for the increased SO₂ removal efficiency at higher pressure are (1) a shift in the vapor-liquid equilibrium or (2) the lower gas velocity through the trays. (At high gas velocities, the bubbles through the sieve trays have more of a tendency to combine and make a channel through

the liquor layer on the trays rather than passing through the liquor as individual bubbles.) The doubling in absolute pressure (from atmospheric to 15 psig pressure) reduces the gas volume and its velocity through the trays to one-half that at atmospheric pressure.

To test the hypothesis that lower gas velocity was contributing to increased tray efficiency, pressure Runs DA22 and 23 were made. Run DA23 was made under all the same conditions as Run DA22 (3% inlet SO₂, Quench L/G = 24, Tray L/G = 38, 15 psig pressure) except for tray gas velocity. From these two runs at the same pressure (DA22 and DA23), it is apparent that the lower gas velocity in Run DA23 is responsible for a significant portion of the increase in absorber efficiency. If Run DA22 is compared instead with Run DA13, the lower gas velocity appears to account for most of the increase in absorber efficiency. Increasing the operating pressure further to 29 psig in Run DA24 (which reduces gas velocity to about 3.7 ft/sec) improves the absorber efficiency compared to that obtained in Run DA23, but gives about the same efficiency as was obtained in Run DA13 at 15 psig.

From these pressure test runs, it is clear that pressure operation improves absorber efficiency and that a significant part of that improvement is due to reduced gas velocity. It is not clear, however, how much improvement was achieved through reduced gas velocity versus a shift in vapor-liquid equilibrium. It is possible that operation under even higher pressure, say 100 psig, would help resolve this question, but that is well beyond the capability of the current bench-scale test system.

6.2 Initial Analysis of Spray Dryer Testing

Table 6-6 summarizes the results for the bench-scale testing of the spray dryer system and Figures 6-1 through 6-3 are plots of removal efficiency vs. alkali/SO₂ ratio. In these figures, the effect of a lower dryer outlet temperature (closer approach to saturation temperature) is also noted. The main effect of a higher spray dryer inlet temperature is increased evaporative capacity which allows higher alkali-to-SO₂ ratios.

The sulfur dioxide removal efficiencies using hydrated lime agree quite well with data reported by Blythe, et al.¹ In their testing, for example, about 55% removal efficiency was obtained at an alkali-to-sulfur dioxide ratio of 1.4. There does not seem to be any significant difference in sulfur dioxide removal efficiency between the levels achieved at a sulfur dioxide inlet concentration of 10,000 ppm in the bench-scale testing vs. the 1000 ppm inlet concentration used by Blythe, et al. (neither loss in efficiency because of mass transfer rate limitations nor increased efficiency due to a greater driving force from the higher sulfur dioxide concentration). We would, therefore, not expect a significant change by increasing the inlet concentration from 10,000 ppm to 15,000 ppm.

The SO₂ removal efficiencies we obtained in the bench-scale, 31-inch diameter Niro Atomizer spray dryer with soda ash (76

¹Blythe, G.M. and R.G. Rhudy, EPRI Spray Drying Pilot Plant Status and Results," EPA/EPRI Symposium, Hollywood, FL, May 1982.

to 91% efficiency) were higher than those obtained by Gehri², et al., in their 7-ft diameter pilot-scale spray dryer (50 to 60% efficiency). This may be due to the greater residence time in our system (40 seconds vs. about 20 seconds in their system).

In a full-scale system where a filter baghouse would be used in place of the cyclone particulate collector that was used with our (bench-scale) pilot plant unit, a further reduction in sulfur dioxide is usually obtained in the downstream particulate collection system. For example, in full-scale commercial FGD systems utilizing a spray dryer with freshly slaked lime and bag filter particulate collection, about 60 to 70% of the inlet sulfur dioxide is removed in the spray dryer and 10 to 25% in the baghouse to yield overall efficiencies of up to 95%.^{3,4}

6.3 Regression Analysis of Bench-Scale Dual-Alkali Absorber Tests

A preliminary scan of the data presented in Tables 6-3 through 6-5 shows that the apparent tray efficiencies (defined as the drop in gas-phase SO₂ concentration across the tray divided by the gas-phase SO₂ concentration entering

²Gehri, D.C. and M.F. Lewis, "Atomization: The Key to Dry Scrubbing at the Coyota Station," EPA/EPRI Symposium, Hollywood, FL, May 1982.

³Donnelly, J.R., et al., "Spray Dryer FGD Experience Joy-Niro Installations," Ninth EPA/EPRI Flue Gas Desulfurization Symposium, Cincinnati, OH, June 1985.

⁴Robards, R.F., et al., "High Sulfur Spray Dryer Evaluations," Ninth EPA/EPRI Flue Gas Desulfurization Symposium, Cincinnati, OH, June 1985.

the tray) were remarkably constant from tray to tray in any given run. (This is the situation that would be expected if one uses a stoichiometric excess of alkali reagent, as was the case here, resulting in essentially zero equilibrium back-pressure of SO_2 on all trays.) This same examination also showed that the percent of SO_2 removal in the quench was small compared with that occurring in the trays; it was decided for simplicity to lump the removal occurring in the quench with that occurring in the first tray at the bottom of the column.

It was, furthermore, assumed that tray efficiencies were all the same in any one given run, and that the quench and Tray 1 together in effect constituted a single tray. It was then possible to arrive at an average effective tray efficiency for each run, based on the following formula:

$$1 - \text{Overall Eff'y} = (1 - \text{Tray Eff'y})^{(\text{No. of Trays})}$$

or alternatively,

$$1 - \text{Tray Eff'y} = (1 - \text{Overall Eff'y})^{(1/\text{No. of Trays})}$$

Table 6-7 shows the result of these calculations. The first group of columns, marked "Experimental Measurements," shows the number of trays, the gas and liquid flows, the Na_2SO_3 concentration in the liquid, the pressure, and the inlet and outlet SO_2 concentrations. The second group of columns, marked "Tray Efficiencies," contains as its first column the result of applying the above equation to the data. The numbers in this column express the result of each run in terms of an average tray efficiency, "Obsvd E," expressed as a fraction. The values range from 55.0% to 85.7%. Note that they do not seem to be significantly different for the four runs (Runs 7, 8, 8A, and 9) in which the Na_2SO_3 strength was

TABLE 6-7

DEGREE OF FIT OF TRAY-COLUMN DATA ON SO₂ SCRUBBING WITH Na₂SO₃ SOLUTION

Run No.	Experimental Measurements									Tray Efficiency			Outlet SO ₂ Conc'n			SO ₂ Removal Eff'y		
	No. Trays	G scfm	P atm	G/(PA) ft/m	Liquor Conc'n Normlty	L gpm	L/A gpm/ft ²	Initial SO ₂ ppm	Final SO ₂ ppm	Obsvd E	Pred E*	(P-O)/O Diff Pct	Obsvd SO ₂ ppm	Pred SO ₂ ppm	(P-O) Diff ppm	Obsvd Rmvl Pct	Pred Rmvl Pct	(P-O) Diff Pct
1	6	10	1	50.93	0.5	0.38	1.94	33,000	30	0.689	0.687	-0.2	30	31	1	99.91	99.91	0.00
2	6	10	1	50.93	0.5	0.77	3.92	63,000	70	0.678	0.708	4.5	70	39	-31	99.89	99.94	0.05
3	6	10	1	50.93	0.5	1.35	6.88	120,000	65	0.714	0.724	1.4	65	53	-12	99.95	99.96	0.01
4	6	10	1	50.93	0.5	0.38	1.94	31,000	25	0.595	0.687	-1.1	25	29	4	99.92	99.91	-0.01
5	6	10	1	50.93	0.5	0.77	3.92	62,000	30	0.720	0.708	-1.6	30	38	8	99.95	99.94	-0.01
6	6	10	1	50.93	0.5	1.35	6.88	122,000	40	0.737	0.724	-1.8	40	54	14	99.97	99.96	-0.01
7	6	10	1	50.93	1.0	0.38	1.94	31,000	20	0.706	0.687	-2.7	20	29	9	99.94	99.91	-0.03
8	6	10	1	50.93	1.0	0.48	2.44	69,000	20	0.743	0.694	-6.5	20	56	36	99.97	99.92	-0.05
8A	6	10	1	50.93	1.0	0.77	3.92	71,000	30	0.726	0.708	-2.4	30	44	14	99.96	99.94	-0.02
9	6	10	1	50.93	1.0	1.20	6.11	142,000	65	0.722	0.721	-0.2	65	67	2	99.95	99.95	0.00
10	4	10	1	50.93	0.5	0.38	1.94	32,000	150	0.738	0.687	-6.9	150	307	157	99.53	99.04	-0.49
11	4	10	1	50.93	0.5	0.77	3.92	64,000	640	0.684	0.708	3.6	640	463	-177	99.00	99.28	0.28
12	4	10	1	50.93	0.5	1.35	6.88	119,000	820	0.712	0.724	1.7	820	687	-133	99.31	99.42	0.11
13	6	10	2	25.46	0.5	0.38	1.94	28,000	1	0.819	0.794	-3.1	1	2	1	100.00	99.99	0.00
14	6	10	2	25.46	0.5	0.77	3.92	62,000	1	0.841	0.808	-4.0	1	3	2	100.00	99.99	0.00
15	6	10	2	25.46	0.5	1.35	6.88	115,000	1	0.857	0.818	-4.5	1	4	3	100.00	100.00	0.00
16	4	10	2	25.46	0.5	0.38	1.94	32,000	90	0.770	0.794	3.1	90	58	-32	99.72	99.82	0.10
17	4	10	2	25.46	0.5	0.77	3.92	66,000	90	0.808	0.808	0.0	90	90	0	99.86	99.86	0.00
18	4	10	2	25.46	0.5	1.35	6.88	115,000	90	0.833	0.818	-1.8	90	126	36	99.92	99.89	-0.03
19	6	20	1	101.86	0.5	0.76	3.87	60,000	500	0.550	0.557	1.4	500	451	-49	99.17	99.25	0.08
20	6	15	1	76.39	0.5	0.57	2.90	65,000	300	0.592	0.617	4.2	300	206	-94	99.54	99.68	0.14
21	6	10	1	50.93	0.5	0.38	1.94	62,000	65	0.681	0.687	0.8	65	58	-7	99.90	99.91	0.01
22	6	20	2	50.93	0.5	0.77	3.92	32,000	50	0.659	0.708	7.4	50	20	-30	99.84	99.94	0.09
23	6	10	2	25.46	0.5	0.38	1.94	32,000	10	0.739	0.794	7.3	10	2	-8	99.97	99.99	0.02
24	6	10	3	16.98	0.5	0.48	2.44	28,000	1	0.819	0.842	2.8	1	0	-1	100.00	100.00	0.00
Avg.										0.729	0.729	0.1	128	117	-11	99.81	99.82	0.01

Notes: A = Cross-sectional area of 6-inch column = 0.1963 sq ft

E = Observed Tray Efficiency = $1 - (\text{Final SO}_2 / \text{Initial SO}_2)^{(1/\text{No. of Trays})}$ E* = Predicted Tray Efficiency = $1 - (0.03162 * [G/(PA)]^{0.6} / [L/A]^{0.1})$ Calc'd (Final) SO₂ = Initial SO₂ x $(1 - E^*)^{(\text{No. of Trays})}$

doubled. This is what would be expected if a molar excess of alkali was present on all trays, as was the case for all runs (at least 50% excess in all tests).

Inspection of these tray efficiencies shows that they are higher in the runs with lower gas flows or higher pressures, or, to a lesser extent, in the runs with higher liquid rates. We found it possible to explain 86% of the sum-of-squares variance of these tray efficiencies (about their mean) by fitting the tray efficiencies to the simple equation:

$$E = \text{Predicted Tray Eff'y} = 1 - (0.03162 * [G/(PA)]^{0.6} / [L/A]^{0.1})$$

Other combinations of exponents were tested, but the above equation was found to give the best fit. The results of applying this formula are shown in the next column of the table, marked "Pred E." As the next column shows, the tray efficiencies predicted by this formula agree with those observed, generally within a relative error of 3% and always within a relative error of 7%.

If one uses these predicted tray efficiencies to compute the expected exit SO_2 concentration, by using the simple expression:

$$\text{Final SO}_2 = \text{Initial SO}_2 * (1 - \text{Predicted Tray Eff'y})^{(\text{No. of Trays})}$$

then the next group of columns, marked "Outlet SO_2 Conc'n," compares the predicted outlet SO_2 concentration with that observed. Predicted outlet SO_2 concentration for the 6-tray configuration is generally within 10 ppm of that observed, and that for the 4-tray configuration is generally within 130 ppm. The final group of columns compares these same results

on the basis of observed and predicted overall "SO₂ Removal Eff'y." Note that, more often than not, the predicted removal efficiency is within 0.02% of the observed value.

These correlations can be expected to apply only for this same type of sieve tray, and for liquid and gas rates per unit area and for inlet gas concentrations and pressures similar to those in Table 6-7. It should also be borne in mind that all of these runs were made with at least 50% stoichiometric excess of Na₂SO₃ solution, with inlet strengths ranging from 0.5 to 1.0 N.

6.4 Regression Analysis of Bench-Scale Spray Dryer Tests

Three series of spray-dryer tests were conducted:

LS-1 Series, using hydrated lime slurry at 300 g
Ca(OH)₂/liter,

LS-2 Series, using slaked lime slurry at 300 g
Ca(OH)₂/liter, and

SA Series, using soda ash solution at 300 g
Na₂CO₃/liter.

The pertinent data are shown in Table 6-8. In all cases, complete evaporation of all the water added to the dryer was achieved. The resulting outlet humidities and dew points are shown, as well as the adiabatic saturation temperature which could have been achieved if as much water as could have been evaporated had been added instead.

The SO₂ removal efficiency for both types of lime correlates well with the alkali/SO₂ mole ratio, as shown in the table and in Figure 6-4. The results for the soda ash solution lie on a different line--see Figure 6-5. Extrapolation of the least-squares lines of Figures 6-4 and 6-5 indicates that

TABLE 6-8
SPRAY-DRYING SO₂ ABSORPTION TEST RUNS

Run No.	Alkali Type *	Water Rate cc/min	Alkali Rate cc/min	Total Water Evapd g/min	Gas Rate scfm	Gas Rate g/min	Humdty Ratio g/g	Gas Dew Point Temp deg F	Adiabatic Satn Temp deg F	Gas Inlet Temp deg F	Gas Outlet Temp deg F	Approch to Adiabatic Satn Temp deg F	Alk/SO ₂ Mole Ratio	SO ₂ Removal Effy, %	Approch to Satn Ln Ti-Tas ----- To-Tas	Xfer Units -Ln 100-E ----- 100
LS-1 A	Hydrated Ca(OH) ₂	22	64	86	19.9	667	0.129	135	147	980	273	126	1.02	47.6	1.89	0.65
LS-1 B	" "	13	72	85	19.9	667	0.127	134	147	990	264	117	1.15	52.3	1.97	0.74
LS-1 C	" "	4	81	85	19.9	667	0.127	134	148	1020	248	100	1.30	53.1	2.17	0.76
LS-1 D	" "	4	86	90	19.9	667	0.135	136	148	1010	255	107	1.38	55.6	2.09	0.81
LS-1 E	" "	22	86	108	19.9	667	0.162	142	148	1010	239	91	1.38	59.1	2.25	0.89
LS-2 A	Slaked CaO	0	113	113	19.8	664	0.170	143	148	1020	208	60	1.91	71.1	2.68	1.24
LS-2 B	" "	0	132	132	19.8	664	0.199	148	148	1027	198	50	2.23	75.9	2.87	1.42
LS-2 C	" "	0	151	151	19.8	664	0.227	152	149	1035	180	31	2.55	91.0	3.35	2.41
LS-2 D	" "	0	170	170	19.8	664	0.256	156	149	1040	160	11	2.88	99.3	4.39	4.96
SA-2	Na ₂ CO ₃ Solution	18	86	104	19.8	664	0.157	141	147	1000	222	75	1.13	91.3	2.43	2.44
SA-3 A	" "	0	81	81	19.8	664	0.122	133	146	946	227	81	0.96	87.6	2.29	2.09
SA-3 B	" "	0	72	72	19.8	664	0.108	129	145	923	257	112	0.85	82.6	1.94	1.75
SA-4 A	" "	5	86	91	19.8	664	0.137	136	148	1010	241	93	1.18	89.5	2.23	2.25
SA-4 B	" "	18	77	95	19.8	664	0.143	138	147	1000	253	106	1.05	88.7	2.09	2.18

* Note: In all cases, alkali strength was 300 g/l of Ca(OH)₂ or Na₂CO₃.

FIGURE 6-4

SPRAY DRYER REMOVAL OF SO₂ WITH Ca(OH)₂
EFFICIENCY VS. ALKALI/SO₂ MOLE RATIO

FIGURE 6-5

SPRAY DRYER REMOVAL OF SO₂ WITH Na₂CO₃
EFFICIENCY VS. ALKALI/SO₂ MOLE RATIO

close to quantitative removal of SO_2 could be expected at stoichiometries of about 3 moles of lime per mole of SO_2 or about 1.6 moles of soda ash per mole of SO_2 .

Table 6-8 also presents, in the right-hand two columns, an attempt to correlate the number of SO_2 transfer units with a measure of the approach of the dryer outlet gas to the adiabatic saturation temperature. The same parameters are used as in Blythe, et al.⁵ That is:

$$\text{No. of Transfer Units} = -\text{Ln} [(100 - E)/100],$$

where E is the percentage removal of SO_2 in the dryer, and

$$\text{Approach to Saturation} = \text{Ln} [(T_{\text{in}} - T_{\text{as}})/(T_{\text{out}} - T_{\text{as}})],$$

where T_{in} , T_{out} , and T_{as} are respectively the inlet, outlet, and adiabatic saturation temperatures of the gas stream.

The results of this treatment of the data have been plotted in Figures 6-6 and 6-7. The correlations plotted show a decidedly non-linear relation in Figure 6-6 for the lime slurry runs, and a general trend in Figure 6-7 but over a very limited range of saturations. In general, correlations of this type are highly dependent upon the strength of the solution or slurry being added to the spray dryer.

In general, then, while the transfer-unit vs. approach-to-saturation correlation did not agree with the relations in the above paper, the data for SO_2 removal vs.

⁵Ibid. Ref. (1)

FIGURE 6-6

SPRAY DRYER REMOVAL OF SO_2 WITH $Ca(OH)_2$
 SO_2 TRANSFER UNITS VS. WATER SATURATION

FIGURE 6-7

SPRAY DRYER REMOVAL OF SO_2 WITH NaCO_3
 SO_2 TRANSFER UNITS VS. WATER SATURATION

alkali/SO₂ ratio agreed well with the data points shown in that paper for hydrated and slaked lime.

7.0 COMMERCIAL DUAL-ALKALI AND SPRAY DRYER DESIGN FOR ZINC-FERRITE OFFGAS TREATMENT

7.1 Dual-Alkali System

Figure 7-1 shows a general schematic and basic material balance for the application of the dual-alkali process to desulfurization of regeneration offgas from a zinc-ferrite absorber servicing a 100 MWe coal gasifier. Figure 7-2 is a more detailed diagram of the sulfur dioxide absorber unit.

Heat recovery would be practiced both for reheat of the recycle stream and to lower the inlet temperature to the absorber.

The absorber was sized based on the equations developed in Section 6.0 of this report. The forced oxidation reactor design and size were based on literature data. The rest of the process equipment (e.g., reactors, tanks, thickener and filter, etc.) was sized based on the design of the full-scale dual-alkali system at the Louisville Gas & Electric (LG&E) Cane Run Station.⁶ The reasons for using this (full-scale) LG&E system as the basis for the design of a commercial zinc-ferrite offgas treatment process are several. This FGD system is a proven commercial process. The technical and cost data for this system are readily available in the published literature (Ref. 6). Finally, the author of this report is personally familiar with the technical design and cost basis details of this commercial facility at LG&E.

⁶Van Ness, Robert, Lawrence Woodland and Elizabeth Gibson, "Full-Dual-Alkali FGD Demonstration at Louisville Gas & Electric Company," PB83-231364, EPA-600/7-83-039, August 1983.

FIGURE 7-1: PROCESS BLOCK FLOW DIAGRAM FOR DUAL-ALKALI DESULFURIZATION OF ZINC-FERRITE REGENERATION OFFGAS

FIGURE 7-2: DUAL-ALKALI ABSORBER TOWER

Several changes were made in the design of the dual-alkali system for this particular application, however. First, the L/G ratio (gpm per 1000 scfm) would be much higher in this application (60 for this unit vs six for the LG&E unit) because of the much higher sulfur dioxide concentration (40,000 ppm vs 4000 ppm sulfur dioxide). (As an alternative to the higher L/G, the concentration of sodium sulfite could be increased, but the reduction in pump size and cost would be more than offset by an increase in the settler/thickener size to compensate for the slower settling rate in the more concentrated scrubber solution.)

The second difference in the design would be to split the (sodium bisulfite) absorber bleed stream with about one-half routed to the oxidation and regeneration reactors, while the other half was sent directly to the overflow tank. The sodium bisulfite from the absorber would react with the sodium hydroxide stream from the thickener/settler to produce a sodium sulfite solution in the overflow tank for recycle to the absorber. The sodium bisulfite stream entering the oxidation reactor would be converted from bisulfite to bisulfate according to the following equation:

As the bisulfite is oxidized to bisulfate the solution becomes more acidic. The pH could be maintained at the optimum level for oxidation in this application, however, by adding back sodium hydroxide solution from the thickener or sodium sulfite solution from the thickener overflow tank.

The gypsum would be settled out of the slurry from the reaction tank in the settler/thickener. Clarified sodium hydroxide solution would overflow into the overflow tank, while the thickened underflow slurry (gypsum solids) would be

pumped to the filter feed tank at about 25 percent suspended solids.

If the evaporation rate from the thickener clarifier were not sufficient to maintain the water balance, the hydrated lime could be directly introduced to the regeneration reactor instead of slurrying it first, thus reducing or eliminating this water input.

In the overflow tank, the regenerated sodium hydroxide solution would react with the recycled sodium bisulfite solution to generate sodium sulfite solution for return to the absorber according to the following equation:

As an alternative, the sodium hydroxide solution could be recycled directly to the absorber and the sodium bisulfite stream from the absorber to the thickener overflow tank eliminated. This would reduce the system size slightly, but would also reduce the flexibility of the system to deal with system upsets and complicate pH control in the oxidation step. It could also lead to tray pluggage in the absorber by precipitated calcium salts at the higher pH.

The internal recycle to the quench sprays of the absorber shown in Figure 7-2 is necessary to maintain a low pH in the bleed stream to the oxidation reactor.

7.2 Spray Dryer System

A schematic of the proposed spray dryer desulfurization system is shown in Figure 7-3.

Basis: 100 MWe Coal Gasifier

FIGURE 7-3: PROCESS BLOCK FLOW DIAGRAM FOR SPRAY DRYER
 DESULFURIZATION OF ZINC-FERRITE REGENERATION OFFGAS

In this somewhat simpler system, lime slurry is sprayed into the hot gas stream. The slurry water is evaporated and the lime reacts directly with the sulfur dioxide to form calcium sulfite as shown in the following equation:

The calcium solids are collected in a filter baghouse. About 40 percent of the solids would be recycled back to the spray dryer to improve lime utilization. In our bench-scale spray dryer testing, it was found necessary to use approximately three moles of lime per mole of sulfur dioxide to achieve less than 100 ppm in the spray dryer outlet. It is assumed, therefore, that a recycle stream in our proposed system about equal in weight to the fresh lime input (which would provide a 2.7 to one unreacted lime-to-sulfur dioxide ratio) and the added removal capacity in the baghouse would result in a spray dryer system outlet concentration of less than 100 ppm with an inlet concentration of 15,000 ppm.

It should be noted that this type of desulfurization system would be limited to treating zinc-ferrite regeneration offgas having sulfur dioxide concentrations of 1.5 percent SO_2 or less because of feed slurry water evaporation capacity limitation. It should likewise be noted that the volume of offgas from the zinc-ferrite reactor regeneration step in this case would have to be considerably greater than that for a wet scrubbing system because of the lower sulfur dioxide concentration. (For a given weight of coal and sulfur input rate, the total volume of regeneration gas is inversely proportional to the sulfur dioxide concentration.)

In the spray dryer system used as the basis for estimation of capital and operating costs in EPRI Report CS-3342 (December 1983), the spray dryer waste solids combined with fly ash

(without oxidation to sulfate) were considered environmentally acceptable for direct landfilling. If it was determined that these calcium sulfite/hydroxide waste solids were not environmentally suitable for direct landfilling, they would have to be first slurried and then oxidized to the sulfate before disposal and this step would add to the cost for waste disposal.

8.0 ECONOMIC COMPARISONS OF DUAL-ALKALI, SPRAY DRYER AND WELLMAN-LORD TREATMENT OF ZINC-FERRITE REACTOR REGENERATION OFFGAS

8.1 Economic Evaluation of the Dual-Alkali Desulfurization System

The block flow diagram 7-1 (in Section 7.0 of this report) was used as the basis for calculating capital and operating costs for the dual-alkali system as applied to desulfurization of zinc-ferrite absorber regeneration offgas.

The data obtained in the bench-scale testing program was used in the design of the absorber, while the design for the rest of the dual-alkali system was based on the commercial unit constructed at Louisville Gas & Electric Co.'s (LG&E's) Cane Run Station.⁷ The dual-alkali FGD system on LG&E's Cane Run Unit 6 uses two absorbers, 45 ft in height x 32 ft in diameter, to handle a total flue gas flow of up to 1,065,000 acfm (at full load and 300°F) from twin electrostatic precipitators on a 300 MW boiler. The boiler is fired with pulverized coal with a heating value of 11,000 Btu/lb and a sulfur content of up to 5.0%.

The actual equipment and materials cost for the dual-alkali system constructed at LG&E and a breakdown of the actual total capital cost, as constructed in 1978, are given in Tables 25 and 26 of the above literature reference.

Table 8-1 lists our estimate of the equipment (and cost in \$1988) for the dual alkali system shown in Figure 7-1. Table

⁷Ibid Ref. (6).

TABLE 8-1

ESTIMATED FIXED CAPITAL COST FOR DUAL-ALKALI TREATMENT
OF ZINC-FERRITE OFFGAS

Basis: 100 MWe Coal Gasifier, 4% SO₂ (vol.) Offgas, Illinois #6 Coal

<u>Equipment</u>	<u>Size</u>	<u>Estimated Cost \$ July 1988</u>
Heat Exchanger	8,360 sq ft	\$ 230,000
Booster Fan	15,000 scfm @ 15 in W.C.	45,000
Absorber	14 ft diam x 40 ft ht	
	8 trays + quench	150,000
Oxidation Reactor	12,000 gal, lined, agitated	78,000
Air Compressor	2,700 scfm at 10 psig	52,000
Lime Reactor	18,000 gal, agitated, C.S., lined	100,000
Lime Slurry Tank	3,000 gal, agitated C.S.	23,000
Lime Silo	3,400 cu ft	50,000
Settler/Thickener	44 ft diam x 17 ft ht	140,000
Thickener Overflow Tank	52,000 gal, C.S., lined	50,000
Rotary Vacuum Filter	8 ft diam x 10 ft length, 304SS	473,000
Filter Feed Tank	3,000 gal, agitated C.S.	23,000
Sodium Hydroxide Storage	6,000 gal, C.S.	7,000
Lime Weigh Feeder	7,400 lbs/hr	80,000
Linings		100,000
Pumps		
Absorber Feed	12,000	
Absorber Bleed	9,900	
Absorber Quench	7,100	
Thickener Underflow	4,500	
Makeup Caustic	<u>1,500</u>	<u>35,000</u>
	TOTAL FGD SYSTEM EQUIPMENT	\$1,636,000
	Waste Disposal System Equipment	<u>\$1,680,000</u>
	TOTAL PROCESS EQUIPMENT	\$3,316,000
Process Instrumentation (10% Process Equipment)		332,000
Process Piping/Insulation/Heat Tracing (15% Process Equipment)		497,000
Electrical (5% Process Equipment)		166,000
Foundations & Buildings (20% Process Equipment)		664,000
Auxiliaries (5% Process Equipment)		<u>166,000</u>
	TOTAL MATERIALS COST	\$5,141,000
Erection Costs (Direct Labor, Overhead & Field Supervision)		<u>2,056,000</u>
	TOTAL PROCESS CAPITAL	\$7,197,000
Engineering Costs (15% of TPC)		1,079,000
Allowance for Funds during Construction & Startup (5.2% of TPC)		<u>374,000</u>
	TOTAL CAPITAL REQUIREMENT	\$8,650,000

8-2 lists the power requirements for this system, and Table 8-3, the estimated annual operating cost.

In arriving at the total plant investment, we have used the actual cost factors experience during the construction of LG&E's Cane Run unit. These costs also include such items as pre-operation and initial chemical costs. These actual capital investment costs (\$/KW) are substantially lower than those recently estimated by EPRI (EPRI CS-3342 December 1983 vol 1) for a dual-alkali process based on the FMC dual-alkali process.

Unit costs for Table 8-3 were taken from a more recent EPRI report (EPRI-P-4463-SR) where available, but were supplemented with updated costs from EPRI-CS-3342 and chemical prices from the September 1988 Chemical Marketing Reporter. Some of these unit costs from the EPRI report which we have used in our cost estimates were considerably higher than those actually experienced by LG&E. (For example, the cost of sludge disposal at LG&E is only about \$3 per ton, and the cost for power less than \$0.03/Kwh.) We have assumed the use of 50 percent caustic soda solution on this smaller scale as an alternative to the soda ash used at LG&E, since it would be much more convenient to handle and would not add significantly to operating cost. The estimated operating cost components for the dual-alkali system are grouped in roughly the same manner as the operating cost components shown for the other two processes in the 1983 EPRI report in order to allow the three alternative processes to be more easily compared. In the EPRI report, Fixed and Variable operating costs are listed in one table (as operating costs), while the capital related-costs are presented in a second table (as part of the levelized busbar cost). For simplicity, our total estimated operating costs,

TABLE 8-2

UTILITY REQUIREMENTS FOR DUAL-ALKALI ZINC-FERRITE OFFGAS TREATMENT

Basis: 100 MWe Coal Gasifier

1. POWER

<u>Equipment</u>	<u>HP</u>
Booster Fan	50
Oxidation Compressor	100
Oxidation Reactor Agitator	15
Lime Reactor Agitator	15
Lime Slurry Tank	5
Lime Feeder	2
Settler/Thickener	5
Rotary Vacuum Filter	30
Absorber Feed	75
Absorber Bleed	40
Absorber Quench	30
Thickener Underflow	20
Caustic Makeup	<u>1</u>
TOTAL	$388 \text{ HP} \times \frac{0.7455 \text{ Kw}}{\text{HP hr}} \times \frac{1}{0.75 \text{ Eff.}} = \frac{400 \text{ kWh}}{\text{hr}}$

$$24 \text{ hrs/day} \times 240 \text{ days/yr} \times 400 \text{ kWh/hr} = 2.304 \times 10^6 \text{ kWh/yr}$$

2. WATER

Lime Slurry 44 gpm

Filter Wash 40 gpm

$$84 \text{ gpm} \times \frac{60 \text{ min}}{\text{hr}} \times \frac{24 \text{ hrs}}{\text{day}} \times \frac{240 \text{ days}}{\text{yr}} = 29,114 \times 10^3 \text{ gal/yr}$$

TABLE 8-3

DUAL-ALKALI DESULFURIZATION SYSTEM OPERATING COST

Basis: 240 days/yr; 24 hrs/day; (0.65 Capacity Factor); \$8.65 Million
 Total Capital Requirement (TCR); 100 MWe Coal Gasifier;
 Illinois #6 Coal (4.3% S dry basis; 11,240 Btu/lb)

	<u>Units/Year</u>	<u>\$/Unit</u>	<u>\$/Year (1988)</u>
<u>Fixed Operating Costs</u>			
Operating Labor	34,880	19.70/m-h	\$ 687,100
Maintenance Labor	11,890	22.00/m-h	261,600
Maintenance Material (2.5% TCR)	--	--	216,200
Supervision	8,000	30.00/m-h	240,000
Plant Engineer/Mgr.	--	--	60,000
Laboratory Support	--	--	40,000
Plant Overhead (30% O&M Labor)			<u>280,600</u>
	FIXED COSTS		\$1,785,500
<u>Variable Operating Costs</u>			
Hydrated Lime	21,220	66.00/ton	1,400,500
Caustic Soda	491	230.00/ton	112,900
Electricity	2,304,000	0.06/kWh	138,200
Process Water	29,114	0.60/1000 gal	17,500
Sludge Disposal	124,420	8.00/ton	<u>995,500</u>
	VARIABLE COSTS		\$2,664,600
<u>Indirect Costs*</u>			
Interest (8.5% of TCR)			735,200
Depreciation (5% of TCR)			432,500
Local Taxes & Insurance (2% OF TCR)			<u>173,000</u>
	TOTAL OPERATING COST		\$5,790,800
	Mills per kWh**		10.2

* Total compares with approximately 15.2% of TCR used in Dec 83 EPRI report.

** $0.65 \times 8760 \text{ hrs} = 5694 \text{ hrs} \times (100,000 \text{ kW} - 400 \text{ kW}) = 5.67 \times 10^8 \text{ kWh/yr}$

including capital related costs, are presented in one table for each of the alternative desulfurization systems.

For the dual-alkali system, the total capital-related (Indirect) costs were estimated at approximately 15.5 percent of Total Capital Requirement (TCR). The capital-related (Indirect) costs presented in the 1983 EPRI report for the FMC dual-alkali process was 15.2 percent of TCR, and for the spray dryer and Wellman-Lord processes, it was 15.4 percent of TCR.

8.2 Economic Evaluation of the Spray Dryer Desulfurization System

Table 8-4 is an estimate for the capital investment for a spray dryer desulfurization system to treat the zinc-ferrite regeneration offgas from a 100 MWe gasifier. This capital investment is based on the flow diagram shown in Figure 7-3. Equipment costs presented for these processes in the December 1983 EPRI report were scaled down by appropriate factors to the size required for the 100 MWe gasifier. As noted earlier in section 7.0 of this report, because the spray dryer desulfurization system is limited to an inlet concentration of about 1.5 percent SO_2 , it would have to treat a larger volume of offgas from the zinc-ferrite reactor than would the dual-alkali absorber, and, therefore, would have higher equipment costs associated with handling the larger gas stream (i.e., higher fan, spray dryer, and bag filter costs). Estimated operating costs for the spray dryer desulfurization system (based on the scaled-down EPRI system) are presented in Table 8-5.

TABLE 8-4

ESTIMATED FIXED CAPITAL COST FOR SPRAY DRYER
DESULFURIZATION OF ZINC-FERRITE REGENERATION OFFGAS

Basis: 100 MWe Coal Gasifier, 1.5% SO₂ (15,000 ppm) and
132,000 acfm @ 1300°F Regeneration Offgas

<u>Equipment</u>	<u>Estimated Cost</u> <u>\$ (July 1988)</u>
SO ₂ Removal System (spray dryer, fabric filter, solids recycle)	\$2,180,000
Flue Gas Handling System (ID/booster fans, ductwork)	1,445,000
Waste Handling System (fly ash blending, trucking)	3,184,000
Reagent Feed System (quicklime storage, ball mill slaker, feed tanks)	<u>2,854,000</u>
TOTAL PROCESS EQUIPMENT	\$9,663,000
TOTAL CAPITAL REQUIREMENT	\$15,122,000

TABLE 8-5

SPRAY DRYER DESULFURIZATION SYSTEM OPERATING COST

Basis: 240 days/yr; 24 hrs/day; (0.65 capacity factor);
 \$15,122,000 Total Capital Requirement (TCR);
 100 MWe Coal Gasifier; Illinois #6 Coal

	<u>Units/Year</u>	<u>\$/Unit</u>	<u>\$/Year</u> <u>(1988)</u>
<u>Fixed Operating Costs</u>			
Operating Labor	32,100	19.70/m-h	\$ 632,400
Maintenance Labor	26,360	22.00/m-h	580,000
Maintenance Material	--	--	865,000
Administration & Support Labor			<u>363,700</u>
	FIXED COSTS		2,441,100
<u>Variable Operating Costs</u>			
Quicklime	32,100	60.00/ton	1,926,000
Electricity	5,600,000	0.06/kWh	330,000
Process Water	60,000	0.60/1000 gal	36,000
Dry Solids Disposal	250,000	8.00/ton	2,000,000
Collector Bags	--	--	<u>30,000</u>
	VARIABLE COSTS		4,322,000
<u>Indirect Costs</u>			
Process Capital Cost (15.4% of TCR)			<u>2,268,000</u>
	TOTAL OPERATING COST		\$9,031,100
$\div 5.5 \times 10^8$ kWh/yr =		Mills/kWh	16.4

TABLE 8-6

ESTIMATED FIXED CAPITAL COST FOR WELLMAN-LORD
DESULFURIZATION OF ZINC-FERRITE OFFGAS

Basis: 100 MWe Coal Gasifier; 4% SO₂ (vol.) Tail Gas; Illinois #6 Coal

<u>Equipment</u>	<u>Estimated Cost</u> <u>\$ July 1988</u>
Heat Exchanger	\$ 230,000
SO ₂ Absorber	150,000
Scrubbing Liquor Tank	50,000
Booster/ID Fan	45,000
Reagent Feed System (Caustic Soda Storage + Pump)	8,500
Regeneration System	12,137,000
By-Product System	3,494,000
Waste Handling System	<u>1,000,000</u>
TOTAL PROCESS EQUIPMENT	\$17,114,500
TOTAL CAPITAL REQUIREMENT	\$27,040,000

TABLE 8-7

WELLMAN-LORD DESULFURIZATION SYSTEM OPERATING COST

Basis: 240 days/yr; 24 hrs/day; (0.65 capacity factor);
 \$27,040,000 Total Capital Requirement (TCR);
 100 MWe Coal Gasifier; Illinois #6 Coal

	<u>Units/Year</u>	<u>\$/Unit</u>	<u>\$/Year</u> <u>(1988)</u>
<u>Fixed Operating Costs</u>			
Operating Labor	80,000	19.70 m-h	\$1,576,000
Maintenance Labor	50,000	22.00 m-h	1,100,000
Maintenance Materials	--	--	820,000
Administration & Support Labor (30% of O&M Labor)	--	--	<u>800,000</u>
FIXED COSTS			\$4,296,000
<u>Variable Operating Costs</u>			
Sodium Hydroxide	500	230.00/ton	115,000
Methane	117,000	8.40/1000 c.f.	983,000
Raw Water	30,000	0.60/1000 gal	18,000
Cooling Water	970,000	0.20/1000 gal	194,000
Low Pressure Steam	202,000	5.00/1000 lbs	1,010,000
High Pressure Steam	102,700	5.50/1000 lbs	565,000
Power	8,130,000	0.06/kWh	488,000
Dry Solids Disposal	28,040	8.00/ton	224,000
Catalyst (Allied)	4,000	1.00/lb	4,000
Catalyst (Claus)	1,200	0.41/lb	<u>500</u>
VARIABLE COSTS			\$3,601,500
<u>Indirect Costs</u>			
Process Capital Cost (15.4% of TCR)			<u>\$4,164,200</u>
TOTAL OPERATING COST			\$12,061,700
Sulfur Credit	7,200	100.00/Long Ton	<u>720,000</u>
NET OPERATING COST			\$11,341,700
÷ 5.3 x 10 ⁸ kWh/yr =		Mills/kWh (Net Operating Cost)	21.4

8.3 Economic Comparison with the Wellman-Lord Sulfur Recovery Process

Finally, estimated capital and operating costs are presented in Tables 8-6 and 8-7 for a sulfur recovery process (Wellman-Lord) that could be applied to the treatment of zinc-ferrite regeneration offgas.

In the Wellman-Lord FGD sulfur recovery process, the sulfur dioxide is first absorbed in sodium sulfite solution to produce sodium bisulfite in the same manner as in the dual-alkali process. The next step in the Wellman-Lord process, however, is recovery of the sulfur dioxide during the evaporation and crystallization of sodium sulfite from the bisulfite solution (instead of reaction to waste calcium salts as in the dual-alkali process) with recycle of the sodium sulfite back to the absorber. Part of the sulfur dioxide generated in the sodium sulfite crystallizer is converted (via the Allied Chemical Process) to H_2S using methane. This H_2S is reacted with the balance of the sulfur dioxide in a Claus reactor to produce elemental sulfur.

Because of the complexity of this multi-step chemical process, the capital cost of this process is much higher than that for either the dual-alkali or spray drying desulfurization process.

The operating cost for the Wellman-Lord process is likewise higher than that for either the dual-alkali or spray dryer desulfurization system (despite the sulfur credit) because of much higher labor requirements and capital investment related costs.

9.0 RESULTS

9.1 Dual-Alkali Bench-Scale Test Results

In the bench-scale testing system it was determined that tray efficiency in a multi-tray absorption system will vary from 60-80% (an average of about 70%) at near atmospheric pressure and a tower velocity of 50 ft/min to 70-90% (average of about 80%) at 15 psig and a tower velocity of 25 ft/min. Other variables such as pH and higher alkali concentrations (>60% excess) seemed to have some but much less effect (1-2%) on SO₂ removal efficiency. (Refer to Section 6.0 of this report.)

In this (standard design) sieve tray absorber system, almost any desired removal efficiency can be achieved by varying the number of transfer stages (trays). Individual tray efficiency (and overall efficiency) can also be increased by reducing gas velocity (or increasing operating pressure) until tray weepage just begins, but operation too close to the weepage point would be too unstable.

In the bench-scale testing, we found that six trays operating at about 70% efficiency were sufficient to reduce inlet concentrations of up to 120,000 to less than 100 ppm in the absorber outlet at a 50 ft/min absorber tower gas velocity. One unexpected result was that the whole range of SO₂ concentrations present through the absorber did not seem to have any effect on individual tray efficiencies.

9.2 Spray Dryer Bench-Scale Test Results

Test results of the bench-scale spray dryer desulfurization system were comparable to results obtained by other researchers for alkali to SO₂ ratios of up to 2.0

When the alkali to SO_2 ratio was increased to about 2.9 (with a 5°F approach to saturation), however, an SO_2 removal efficiency of over 99% was achieved for an SO_2 inlet concentration of about 11,000 ppm.

9.3 Conceptual Designs of the Dual-Alkali Desulfurization System

The conceptual design of the absorber for the dual-alkali desulfurization system for a 100 MWe gasifier/zinc-ferrite reactor was based on the results of this bench-scale testing program. The design of the oxidation system is based on literature data for forced oxidation systems, and the balance of the dual-alkali system is based on the operating dual-alkali system at Louisville Gas & Electric's Cane Run Station (Louisville, KY).

9.4 Conceptual Design of the Spray Dryer Desulfurization System

The conceptual design for the spray dryer desulfurization system was based on the system described in EPRI Report CS-3342 (December 1983), modified for the operating conditions necessary for the higher temperature and higher alkali feed rates that would be encountered in treating the zinc-ferrite regeneration offgas. As noted earlier in this report, the spray drying desulfurization system would be limited to treating offgas containing 1.5% or less SO_2 because of water evaporation requirements. Unlike the dual-alkali system, the spray dryer desulfurization would also require a particulate collection device, such as a baghouse or ESP, both for particulate control and its contribution to SO_2 removal.

9.5 Economic Comparison of the Dual-Alkali and Spray Dryer Desulfurization Systems

As more fully described in Section 8.0 of this report, the dual-alkali system has both lower capital and operating costs than either the spray dryer desulfurization system or the Wellman-Lord sulfur recovery system.

The major reasons for this are that the spray dryer desulfurization system requires both a solids recycle system to increase alkali (lime) utilization and a large particulate collection system, while the Wellman-Lord system is a much larger and more complex chemical process.

10.0 CONCLUSIONS

The bench-scale testing of the dual-alkali system has shown that it is possible to achieve the 99.9% SO₂ removal efficiency with this type of system necessary for treatment of the zinc-ferrite offgas containing up to 120,000 ppm SO₂ to below the 100 ppm SO₂ required for recycle of the gas stream to the zinc-ferrite reactor or discharge to the atmosphere.

The bench-scale testing of the spray dryer desulfurization system has demonstrated that it should be possible to reduce an inlet concentration of 10-15,000 ppm (from the zinc-ferrite reactor regeneration step) to below 100 ppm for discharge to the atmosphere. Because of the loss in temperature and high water vapor content, it would not be practical to recycle the spray dryer treated reactor offgas back to the zinc-ferrite reactor. Conceptual designs prepared for both the dual-alkali and spray dryer desulfurization systems (presented in Section 7.0) were used as the basis for estimates of capital and treatment costs for these two processes (Section 8.0).

The estimated capital cost for the dual-alkali treatment system at \$8.6 million for a 100 MWe gasifier/zinc-ferrite desulfurizing system is less than that calculated for an equivalent spray dryer desulfurization system (at about \$15 million) or the Wellman-Lord sulfur recovery process (at about \$27 million). Likewise, the operating cost for the dual-alkali system is estimated at about 10 mills per kWh compared with about 16 mills per kWh for the spray dryer system, and 21 mills per kWh for the Wellman-Lord sulfur recovery process.