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THE EDGEWATER COOLSIDE PROCESS DEMONSTRATION  
A TOPICAL REPORT

by

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## ABSTRACT

The Edgewater Coolside process demonstration met the program objectives which were to determine Coolside SO<sub>2</sub> removal performance, establish short-term process operability, and evaluate the economics of the process versus a limestone wet scrubber. During the process demonstration, the Edgewater Station remained in compliance with all applicable air, water, and solid waste disposal regulations.

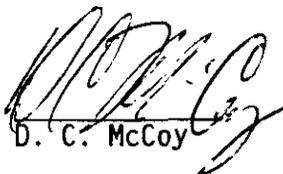
On a flue gas produced from the combustion of 3% sulfur coal, the Coolside process achieved 70% SO<sub>2</sub> removal using commercially-available hydrated lime as the sorbent. The operating conditions were Ca/S mol ratio 2.0, Na/Ca mol ratio 0.2, and 20°F approach to adiabatic saturation temperature ( $\Delta T$ ). During tests using fresh plus recycle sorbent, the recycle sorbent exhibited significant capacity for additional SO<sub>2</sub> removal. The Edgewater Coolside SO<sub>2</sub> removal response to process variables was similar to the Consol pilot-scale results.

The longest steady state operation was eleven days at nominally Ca/S = 2, Na/Ca = 0.22,  $\Delta T$  = 20-22°F, and 70% SO<sub>2</sub> removal. The operability results achieved during the demonstration indicate that with the recommended process modifications, which are discussed in the Coolside process economic analysis, the process could be designed as a reliable system for utility application.

Based on the demonstration program, the Coolside process capital cost for a hypothetical commercial installation was minimized. The optimization consisted of a single, large humidifier, no spare air compressor, no isolation dampers, and a 15 day on-site hydrated lime storage. The levelized costs of the Coolside and the wet limestone scrubbing processes were compared. The Coolside process is generally economically competitive with wet scrubbing for coals containing up to 2.5% sulfur and plants under 350 MW<sub>e</sub>. Site-specific factors such as plant capacity factor, SO<sub>2</sub> emission limit, remaining plant life, retrofit difficulty, and delivered sorbent cost affect the scrubber-Coolside process economic comparison.

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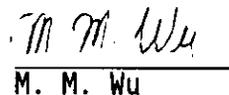
  
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## SECTION 1

### EXECUTIVE SUMMARY

#### INTRODUCTION

The Coolside process is based on injecting a dry sorbent into a humidified flue gas to reduce SO<sub>2</sub> emissions from coal-fired boilers. (A simplified schematic of the Coolside process is shown in Figure 74, page 155 of the text.) Under the Clean Coal Technology Program, the Department of Energy co-funded the 105 MW<sub>e</sub> Edgewater Unit No. 4, Boiler 13 Coolside demonstration. Other project participants were: the Babcock and Wilcox Company (prime contractor and co-funder); the State of Ohio Coal Development Office (co-funder); the Ohio Edison Company (host utility); and CONSOL Inc. (Coolside process developer and co-funder). While not directly involved in the Coolside demonstration, the U.S. Environmental Protection Agency (EPA) made vital contributions to the program. EPA funded the pilot-scale humidifier test program and the design of the Edgewater humidifier. In addition, EPA permitted the use of certain equipment after the EPA LIMB test program concluded. In addition to providing the site, Ohio Edison operated and maintained the Coolside process equipment.

Edgewater Unit No. 4 was retrofitted with a flue gas humidifier between the air preheater exit and the electrostatic precipitator (ESP) inlet and with a hydrated lime bulk storage and feed system. As part of the EPA LIMB demonstration program, Boiler 13 was retrofitted with B&W XCL low NO<sub>x</sub> burners. Due to the short duration of the demonstration program, the existing gunnite-lined steel chimney was not modified for operation at close approach to adiabatic saturation temperature. To protect the chimney from condensate damage, a steam reheater was installed downstream of the ESP. Only operating practice changes were made to the ESP, ash hopper unloading system, ash transport system, ash storage system, and ash silo unloading system.

#### OBJECTIVES

The major objectives of the Coolside demonstration were achieved.

- Demonstrate up to 70% SO<sub>2</sub> removal with high-sulfur coal.
- Demonstrate short-term operability at commercial scale.

- Develop a data base to design a commercial Coolside installation.
- Develop Coolside process economics.

By achieving these objectives, the commercialization of the Coolside process can be accelerated.

#### DESULFURIZATION PERFORMANCE

The Coolside process achieved 70% SO<sub>2</sub> removal using a commercially-available hydrated lime while burning a 2.8-3.0% sulfur coal at Coolside design operating conditions. The design conditions were: 2.0 calcium-to-sulfur mol ratio (Ca/S); 0.2 sodium-to-calcium mol ratio (Na/Ca); and 20°F\* approach to adiabatic saturation temperature ( $\Delta T$ ). Coolside SO<sub>2</sub> removal is a strong function of Ca/S, Na/Ca, and  $\Delta T$ . At Edgewater, the SO<sub>2</sub> removal was not sensitive to coal sulfur content (flue gas SO<sub>2</sub> concentration).

The Coolside process sorbent utilization was about 33% at Coolside design operating conditions. The spent sorbent has significant capacity for additional SO<sub>2</sub> removal. During the Edgewater test program, limited full-scale spent sorbent recycle tests were conducted. The Edgewater results confirmed pilot plant test results that the recycle sorbent has a significant SO<sub>2</sub> removal capacity. At 18°F approach and with neither fresh hydrated lime nor fresh sodium additive feed, and 6800 pph of recycle feed (equivalent to a Ca/S = 0.5), the system SO<sub>2</sub> removal was 22%.

Throughout the Coolside demonstration, the Edgewater Station remained in compliance with all environmental regulations. The Edgewater emission limitations are: 3.4 lb SO<sub>2</sub>/10<sup>6</sup> Btu heat input (30-day rolling average), 0.1 lb particulate matter/10<sup>6</sup> Btu heat input, and 20% opacity (six minute average). There were no NO<sub>x</sub> emission limitations in effect at the time of the demonstration. During the test program with high-sulfur coal, the Coolside process was the SO<sub>2</sub> compliance technology.

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\*For those more familiar with metric units, see the conversion table in Appendix E.

Sorbent selection is an important parameter for the Coolside process. The full-scale results using two commercially-available hydrated limes confirmed the sorbent reactivity differences observed in the pilot-scale sorbent selection study. The lower-reactivity sorbent provided 35% (relative) lower average sorbent utilization than the "best" sorbent. The two commercially-available hydrated limes tested at Edgewater represented the "best" sorbent--cost not considered--and the lowest-delivered-cost hydrated lime.

The full-scale Coolside SO<sub>2</sub> removal response to process variables was similar to the Consol pilot-scale results. This demonstrated that pilot plant testing can be used as a process design tool. The pilot-scale SO<sub>2</sub> removals were correlated as a function of Ca/S, Na/Ca, and  $\Delta T$ . The system SO<sub>2</sub> removals at Edgewater were lower than that observed in pilot plant testing.

#### COOLSIDE OPERABILITY/PERFORMANCE

The Edgewater Coolside demonstration program was designed to establish short-term process operability. Steady state operation was maintained for 11 days at a nominal Ca/S = 2.0, Na/Ca = 0.22,  $\Delta T$  = 20-22°F, and 70% SO<sub>2</sub> removal. The 11-day test established short-term Coolside process operability. Although long-term operability was not established, a reliable Coolside system can be designed for utility application with the process modifications discussed in this report.

Edgewater humidifier operations identified three problem areas: horizontal humidifier floor deposits, internal nozzle deposits, and deposits on the atomizer lances. These problems can be addressed adequately in a commercial design. The horizontal humidifier floor deposits consisted of a fine, dry dust which settled out from the low velocity flue gas and larger fragments of wall deposits. In a commercial unit, either a vertical humidifier or a horizontal humidifier including installation of floor sootblowers and a hopper at the humidifier exit would eliminate the problem. The water atomizer caps showed signs of wear after 4-6 months of operation. In a commercial unit, the nozzles would be hardened or equipped with ceramic inserts to reduce erosion. In addition to hardening and ceramic inserts, water filtration would be improved to remove fine silt, sand, and grit. Atomizer lance deposits would be minimized by injecting the hydrated

lime in the same plane as the atomizer water and including a nozzle cleaning system such as rappers, sootblowers, or brushes.

During the Coolside demonstration, the ESP operations met the plant stack opacity limit (20%). The Edgewater stack opacity was less than 10%--typically less than 5%. However, at high ESP inlet solids loading (>16,000 pph), ESP operations deteriorated after about seven days of operation and Coolside Ca/S and recycle rate were lowered to maintain the opacity limit. The Edgewater ESP design specific collecting area (SCA) is 612 ft<sup>2</sup>/MACF. During Coolside operations, the ESP operations were modified as follows: the plate-rapping frequency was increased on all fields and intermittent energization was used on the first two fields to maintain field energization with limited success. During a scheduled boiler outage, the ESP was inspected. The inspection revealed high-tension wire deposits on the wires which were farthest from the rappers. In a commercial Coolside installation, a higher intensity rapper system is recommended. In a retrofit installation, the utility may wish to consider upgrading the transformer/rectifier sets along with installation of high-emission electrodes and high-tension frame stabilizer bars to optimize ESP performance.

Differences in the collected Coolside ash size distribution and mass flow to the ash storage silo increased the ash silo baghouse cleaning frequency which, in turn, limited the solids throughput. Due to the short duration of the Coolside test program (six months), the ash silo baghouse was not modified to increase throughput. In a commercial Coolside installation, the ESP and ash handling system must be thoroughly evaluated for Coolside operating conditions to eliminate potential bottlenecks.

Throughout the Coolside test program, Ohio Edison continued to operate the B&W XCL-low NO<sub>x</sub> burners. During the EPA program, NO<sub>x</sub> emissions were 0.48 lb/10<sup>6</sup> Btu. The Coolside process did not significantly reduce NO<sub>x</sub> emissions from this emission rate.

#### WASTE MANAGEMENT

During the Coolside demonstration, the solid waste was disposed in a permitted landfill as a nonhazardous solid waste. The Coolside waste was

evaluated using the Environmental Protection Agency Extraction Procedure (EP). The leachates from the EP were within Resource Conservation Recovery Act limits. The concentrations of trace elements, sulfate, and total dissolved solids (TDS) in the leachates were less than thirty times the EPA primary and secondary drinking water standards (Ohio EPA requirements).

Coolside waste has different physical and chemical properties compared to bituminous coal fly ash. The Coolside ash physical properties changed the solids flowability from the ESP hoppers. Fluidizing air was required to improve flow characteristics of the hydrate lime-fly ash mixture. The Coolside waste has a high calcium content, may have an elevated sodium content, and has a lower bulk density. The major components of Coolside waste are fly ash,  $\text{Ca(OH)}_2$ ,  $\text{CaSO}_3$ , and  $\text{CaSO}_4$ . Minor components are  $\text{Na}_2\text{SO}_3$ ,  $\text{Na}_2\text{SO}_4$ , and  $\text{CaCO}_3$ . Coolside waste leachate, like other dry lime-based FGD wastes, is alkaline. Because of the high "free" calcium content, Coolside waste has cementitious properties.

#### COMMERCIAL DESIGN/ECONOMICS

Based on the Coolside  $\text{SO}_2$  removal performance versus Ca/S, Na/S, and  $\Delta T$  and process operability, the Coolside commercial design was developed. The commercial design minimized initial investment cost and annual revenue requirements. The Coolside process capital cost was minimized by: using a single, large humidifier; eliminating the spare air compressor; eliminating isolation dampers; and reducing on-site sorbent and additive storage to 15 days. The commercial design would be typical of the  $n^{\text{th}}$  plant design and parallels the progress shown in utility wet scrubber applications.

The Coolside capital and total levelized annual revenue requirements based on 70%  $\text{SO}_2$  removal were compared to those for wet limestone, forced oxidation scrubbing based on 95%  $\text{SO}_2$  removal. Twelve cases were estimated for each process. Boiler capacities were 100, 150, 250, and 500  $\text{MW}_e(\text{net})$ . For each boiler capacity, the coal sulfur contents were 2.2, 3.7, and 5.2  $\text{lbs SO}_2/10^6 \text{ Btu}$  (nominal 1.5, 2.5, and 3.5% sulfur coals). The optimized Coolside capital cost is 40 to 45% of the corresponding wet limestone FGD capital cost. For example, for the 2.2  $\text{lb SO}_2/10^6 \text{ Btu}$  and 250  $\text{MW}_e(\text{net})$  boiler capacity, the Coolside and wet limestone FGD capital costs are \$74/kW and \$184/kW, respectively.

The Coolside and wet FGD total levelized annual revenue requirements were compared. For the financial bases listed in Table 16 and a 250 MW<sub>e</sub>(net) boiler capacity firing a 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu coal, the total levelized annual revenue requirements for the Coolside process and the wet FGD process are \$567 and \$612 per ton of SO<sub>2</sub> removed, respectively. The comparison of total annual revenue requirements is extremely site-specific and is dependent upon factors such as: remaining plant life, plant capacity factor, cost of sorbent, cost of waste disposal, SO<sub>2</sub> removal requirement and power cost. The Coolside process economics are favored when the required SO<sub>2</sub> removal is moderate, i.e., less than 70%; the plant capacity factor is less than 65%; the hydrated lime cost is less than \$60/ton, and the waste disposal cost is low. The wet FGD process economics are favored with higher sulfur coals (greater than 90% SO<sub>2</sub> removal), plant capacity factor is 65% or greater, plant life is 20 years or longer, and the ratio of hydrated lime to limestone cost is greater than 4, and waste disposal cost is high.

As a rule of thumb, the Coolside process is cost competitive with wet limestone FGD for coals containing up to 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu and boiler capacities below 250-300 MW<sub>e</sub>(net). An important assumption in this analysis is that the reference plants are equipped with a 400 SCA ESP. The collection efficiency for a 400 SCA ESP is adequate to comply with a particulate emission limit of 0.1 lb/10<sup>6</sup> Btu and permit the use of recycle plus fresh sorbent addition. Confident application of the Coolside process to utility units with smaller ESPs would require additional analysis and/or pilot-scale testing. If additional fields must be retrofitted to the existing ESP, the Coolside capital cost would increase dramatically.

## RECOMMENDATIONS

Further process development is justified to widen the applicability of the Coolside process. This was the first full-scale demonstration of the Coolside process. The following recommendations could improve Coolside performance and economics.

1. Demonstration of long-term Coolside process operability is recommended. The demonstration would implement the process recommendations

made in this report and confirm the expected improvements in process operability.

2. Research to improve sorbent utilization is recommended. Hydrated lime is the single largest operating cost for the Coolside process. Improving sorbent utilization would reduce the annual hydrated lime consumption, reduce or eliminate the soda ash consumption, and lower annual waste disposal cost. Improved sorbent utilization could be achieved by operation at closer approach to adiabatic saturation temperature, if process operability can be maintained. Alternatively, an improved sorbent may be developed.
3. Research to improve water atomizer performance is recommended. Developing an atomizer which generates less than 1% large (>100 micron) droplets at atomization air-to-water weight ratios of less than 0.45 would lower energy consumption and capital cost. The atomization air compressor accounts for typically over 80% of the process energy consumption. Lowering the air pressure or volume required to atomize the water would significantly lower parasitic energy requirements. Reducing or eliminating the weight fraction of water droplets greater than 100 microns would result in a smaller, less costly, easier-to-retrofit humidifier.
4. Research is recommended to improve the understanding of ESP collection of spent Coolside sorbent to allow wider process application and recycle operation to improve sorbent utilization. For example, a demonstration is required to confirm that more intense rapping will prevent high-tension wire deposits and that the intense rapping prevents ESP performance deterioration.

## SECTION 2

### INTRODUCTION

Under sponsorship the U.S. Department of Energy, Pittsburgh Energy Technology Center, Office of Clean Coal Technology Program, the Coolside process was demonstrated on the 105 MW<sub>e</sub> Unit No. 4, Boiler 13 at the Ohio Edison Edgewater Power Plant, Lorain, Ohio. In addition to the Department of Energy, other project participants were the State of Ohio Coal Development Office (co-funder), the Babcock and Wilcox Company (prime contractor and co-funder), the Ohio Edison Company (host utility), and CONSOL Inc. (technology developer and co-funder). The Coolside demonstration was conducted between July 1989 and mid-February 1990. The Coolside demonstration was completed prior to the passage of the Clean Air Act Amendments of 1990. The Coolside economics were developed to comply with the Clean Air Act Amendments. It was outside the scope of the project to evaluate the impact of SO<sub>2</sub> emission credits on technology selection.

The Coolside process is a duct sorbent injection process developed by CONSOL Inc. for retrofit SO<sub>2</sub> control on a coal-fired boiler. The objectives of the program were to demonstrate: up to 70% SO<sub>2</sub> emission reduction from the combustion of high-sulfur coal, short-term Coolside process operability, and Coolside capital costs were substantially less than cost of a wet scrubber. The Coolside operability and performance data were used to estimate process capital and levelized operating costs. The boiler fired compliance (nominal 1.4 wt % sulfur) and noncompliance (nominal 3 wt % sulfur) Ohio coals. The demonstration program evaluated the effect of varying Ca/S, Na/Ca, and  $\Delta T$  on SO<sub>2</sub> removal. In addition to process variable effects, the effect of sorbent properties was evaluated in pilot- and full-scale tests. Two commercially available hydrated limes were tested at Edgewater, and twelve sorbents were tested at the Consol pilot plant, Library, Pa. Key process variables were evaluated in short-term (6-8 hour) parametric tests and longer term (1-11 day) process operability tests. Sorbent once-through and recycle process tests (in which a portion of the collected ash from the ash storage silo is reinjected into the flue gas to increase sorbent utilization and lower operating costs) were performed. The

solid waste generated during the demonstration program was disposed in a permitted non-hazardous landfill. The State of Ohio classified the Coolside waste as "solid waste."

Full-scale Coolside tests were conducted at the Edgewater Unit No. 4, which has a turbine nameplate rating of 105,000 kW. Boiler 13 is a Babcock and Wilcox front wall, pulverized coal-fired boiler which has a capacity of 690,000 pph steam (design maximum continuous rating). The boiler was retrofitted with B&W XCL burners for the EPA LIMB demonstration program. The flue gas humidifier was constructed on the boiler house roof between the air preheater and the ESP. The ESP has a design specific collecting area of 612 ft<sup>2</sup>/MACF and was not modified for the test program. In addition to the flue gas humidifier, the Edgewater plant was retrofitted with a hydrated lime storage silo and gravimetric, pneumatic hydrated lime feed system and with a sodium hydroxide storage tank and metering system. Operating procedure changes were required for ash hopper unloading, ESP operations, and ash storage silo unloading.

Prior to the demonstration, pilot-scale tests were conducted to select the commercially-available hydrated limes to be tested and to develop process performance data applicable to the Edgewater site-specific conditions. The pilot plant results were used to develop the demonstration program and for subsequent data interpretation of the full-scale results.

The pilot plant test program was conducted at the 0.1 MW<sub>e</sub> Coolside test facility which is designed to study Coolside desulfurization performance over a wide range of process conditions. The pilot unit allows site-specific simulation of the flue gas conditions, including gas composition, temperature, solids loading, and humidifier residence time. The unit has a 8.3 inch ID x 20 foot long, vertical downflow humidifier installed with a commercial two-fluid atomizer. A pulse-jet baghouse is used for particulate removal. Continuous SO<sub>2</sub> and O<sub>2</sub> analyzers are used for the measurement of SO<sub>2</sub> removal across the humidifier and across the system (humidifier plus baghouse).

## SECTION 3

### DESULFURIZATION PERFORMANCE

#### SUMMARY

The demonstration of the Coolside flue gas desulfurization process at the Edgewater Station was operated continuously at conditions set to maintain compliance. The demonstration included individual parametric tests during which plant operators and test personnel set the test conditions. The reported results reflect nominal variations in coal sulfur content, sorbent purity, and equipment and instrumentation characteristics. Since the demonstration sought to identify expected SO<sub>2</sub> removal performance for a commercial, full-scale system, the discussion of results in this section reflect averages taken over a number of runs and test durations with a variability of less than ten percent. Precise values for specific tests are presented in the accompanying tables and figures.

The Coolside desulfurization process was demonstrated successfully at the 105 MW<sub>e</sub> Ohio Edison Edgewater Unit No. 4 Boiler 13 using coal containing 1.2 to 3.0 wt % sulfur. In once-through tests using hydrated lime A with NaOH addition to the humidification water (0.2 Na/Ca mol ratio), the observed SO<sub>2</sub> removals at a nominal 20°F approach to saturation averaged 70% at 2.0 Ca/S and 45% at 1.0 Ca/S. Lime A was the most active sorbent in the pilot sorbent evaluation tests conducted for the demonstration. In tests without additive, the observed SO<sub>2</sub> removals at a nominal 20°F approach were 35 to 45% at 2.0 Ca/S with the same hydrated lime. At 25°F approach, observed SO<sub>2</sub> removals at given Ca/S and Na/S conditions were lower by 5 to 10% (absolute). These results indicate that the process should be designed for the closest approach conditions possible at a given retrofit installation. The process operability data at Edgewater support the conclusion that humidifier operation at a 20°F approach to adiabatic saturation temperature is feasible.

Edgewater tests of hydrated lime G, the lowest cost sorbent (based on delivered price) tested in the pilot plant, showed SO<sub>2</sub> removals lower by 5-10% (absolute) than for hydrated lime A with 0.2 Na/Ca mol ratio at 25°F approach and comparable Ca/S mol ratio. For tests without sodium additive (0 Na/Ca ratio), the SO<sub>2</sub> removals were lower than comparable tests with hydrated lime A. With hydrated lime G, the target approach to saturation had to be increased to 25°F, because of humidifier operating limitations in the latter part of the demonstration due to atomization nozzle problems (detailed information on the demonstration operation is provided in Section 4 of this report). Hydrated limes A and G have the same designations as used in the pilot support test reports. The differences in the SO<sub>2</sub> removals observed between the two sorbents are consistent with that observed in the pilot support tests.

In addition to the once-through tests, sorbent recycle tests also were performed. The results confirm that recycle sorbent has the capacity for significant additional SO<sub>2</sub> removal. However, because of frequent changes in recycle test conditions, the steady-state recycle process performance could not be determined. The test condition changes resulted from having to operate the humidifier with flue gas bypass during high load (>70 MW<sub>e</sub>) conditions, and because the NaOH pump was out of service during part of the test period due to pump seal problems. Thus, the composition of the recycle sorbent was subject to frequent changes.

The SO<sub>2</sub> removals were calculated from the concentration of SO<sub>2</sub> in the flue gas measured at the humidifier entrance and at the ESP outlet using continuous gas analyzers. The SO<sub>2</sub> measurements were corrected for air in-leakage using continuous O<sub>2</sub> analyzer measurements at both locations. The data from the wet-basis analyzers at the humidifier inlet were corrected for moisture using wet/dry bulb measurements. The ESP outlet gas analyzers were dry-basis. The measured SO<sub>2</sub> removals were confirmed by analyzing spent sorbent samples collected from the ESP.

The system SO<sub>2</sub> removals at Edgewater were somewhat lower than in pilot plant tests at similar Ca/S, Na/Ca, and temperature approach conditions. The difference in the particulate collection devices (ESP at Edgewater versus

baghouse at the pilot plant) was probably responsible for at least part of the difference. Since a baghouse provides more effective gas/solid contact than an ESP, the SO<sub>2</sub> removals at Edgewater were projected to be lower compared with those obtained in the baghouse-equipped 0.1 MW<sub>e</sub> pilot plant. Differences in hydrated lime distribution in the flue gas also may have been partially responsible for the difference in SO<sub>2</sub> removals.

The Coolside process tests were conducted over a 6 1/2-month period. For the last 4 1/2 months, testing was round-the-clock. The longest continuous test period at 20°F approach conditions without a humidifier shutdown was 11 days. A noteworthy accomplishment of the program was that the Coolside process kept SO<sub>2</sub> emissions in compliance during high sulfur coal firing tests.

#### ONCE-THROUGH PROCESS PERFORMANCE RESULTS

##### Effect of Ca/S Ratio Variation

This operating variable was important for controlling the SO<sub>2</sub> removal at a desired level because the removal increased in a predictable manner with an increase in the Ca/S mol ratio for both of the hydrated limes tested. In once-through tests with NaOH addition to the humidification water (0.2 Na/Ca mol ratio) using hydrated lime A, the observed SO<sub>2</sub> removals were 45% at 1.0 Ca/S and averaged 70% at 2.0 Ca/S. In tests without additive, the observed SO<sub>2</sub> removals at a nominal 20°F approach were 35 to 45% at 2.0 Ca/S with hydrated lime A. At 25°F approach, the observed removals at similar Ca/S and Na/Ca ratios were lower by 5-10% (absolute). The sorbent utilization was dependent upon Ca/S ratio and decreased by about 7.5% (absolute) as the Ca/S was increased from 1.0 to 2.0. Hydrated lime A was the most active sorbent in the pilot sorbent evaluation tests conducted for the demonstration.

Using hydrated lime G, the same trend with Ca/S was observed. As expected from pilot plant results, the removals were somewhat lower than with hydrated lime A. SO<sub>2</sub> removals in once-through tests at 25°F approach with NaOH additive (0.2 Na/Ca mol ratio) were 37% at 1.3 Ca/S and 55% at 1.9 Ca/S. For tests without additive, the removals were 29% at 1.4 Ca/S and 34% at 1.9 Ca/S. The

approach to adiabatic saturation was intentionally increased by 5°F to 25°F for tests with hydrated lime G because of a limitation in the humidifier operation.

Figure 1 shows the once-through SO<sub>2</sub> removals using hydrated lime A (with NaOH additive at 0.17 to 0.24 Na/Ca mol ratio and without the additive) plotted against the Ca/S molar ratio for tests at 19 to 22°F approach (circles) and 23 to 27°F approach (crosses) to adiabatic saturation temperature. In the figure, the SO<sub>2</sub> removal data with the additive are grouped separately for the two approach ranges because the SO<sub>2</sub> removal level is sensitive to small changes in the approach to adiabatic saturation. Although there is some variation in SO<sub>2</sub> removals at similar Ca/S ratios, Figure 1 clearly shows the trend of higher SO<sub>2</sub> removals at higher Ca/S ratios. Tests without sodium additive at 23 to 27°F approach showed a similar trend.

Tests without additive also were made using lime A early in the test program (August 28 to September 28, 1989). Since the humidifier inlet gas analyzers were not operational during these early tests, the SO<sub>2</sub> removals were calculated using the stack gas analyzer data collected before and after the tests for the baseline flue gas SO<sub>2</sub> content. Although the set approach was 20°F, the actual approach to saturation varied from 20°F to over 30°F during this test period. Thus, considerably more scatter occurred in the SO<sub>2</sub> removals calculated for this data set (Figure 2). Even with the high degree of scatter, the effect of Ca/S ratio on SO<sub>2</sub> removal was evident and is similar to the no-additive results in Figure 1.

Tests with hydrated lime G showed a similar effect of Ca/S ratio on SO<sub>2</sub> removal. Figure 3 shows the SO<sub>2</sub> removals using lime G plotted against Ca/S molar ratio at 23 to 27°F approach. With the NaOH additive (0.17 to 0.24 Na/Ca mol ratio), the SO<sub>2</sub> removals increased with increasing Ca/S ratio. For tests without additive, the SO<sub>2</sub> removals showed a similar trend with a more gradual slope (Figure 3). No tests were performed using hydrated lime G at 19 to 22°F approach because the humidifier performance had deteriorated to the point where operation at 25°F approach to saturation was necessary. This is explained in more detail in the Process Reliability/Operability section (Section 4) of this report.

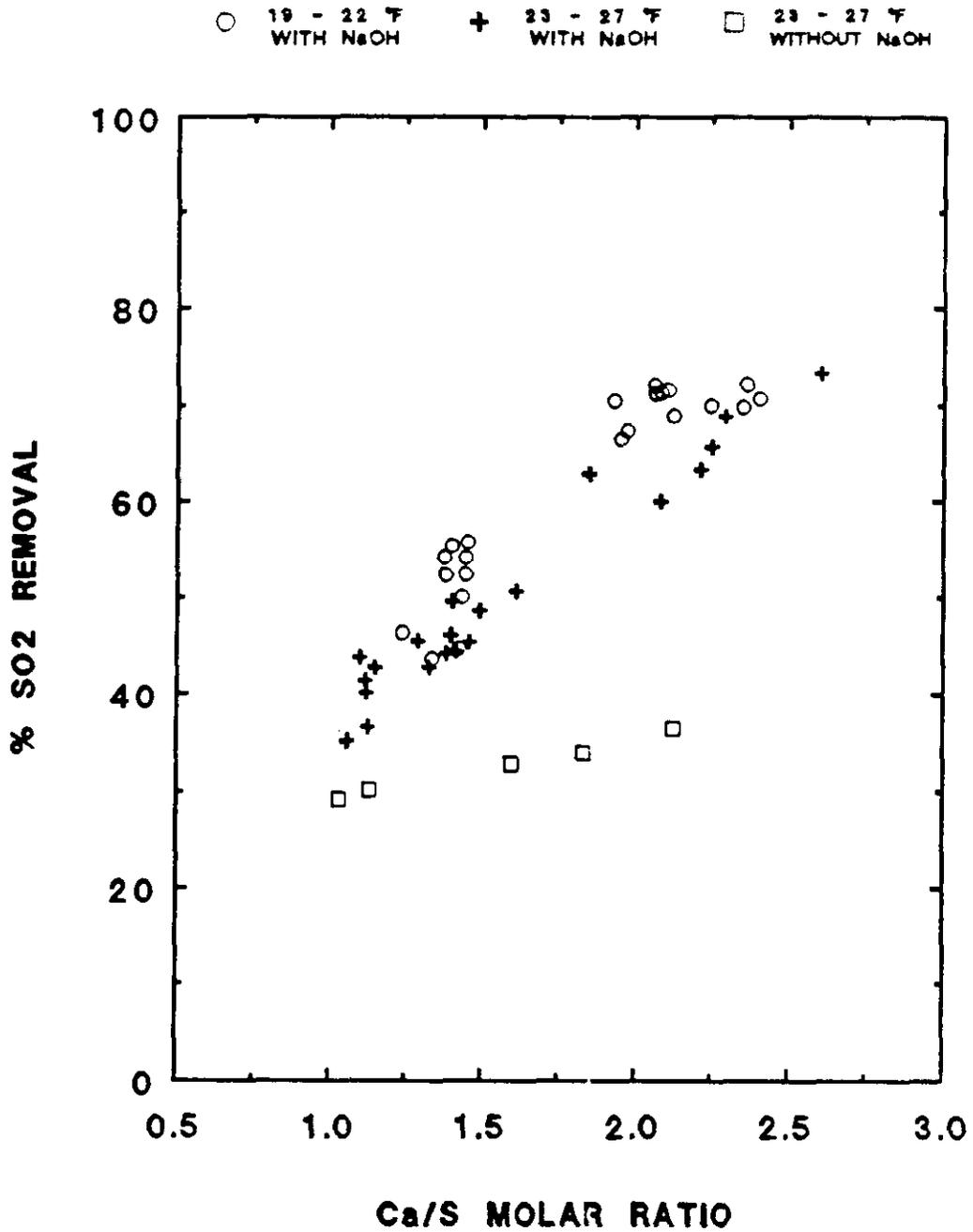


Figure 1. SO<sub>2</sub> removals observed using hydrated lime A. Circles represent 0.17 to 0.24 Na/Ca and 19 to 22°F approach to adiabatic saturation; crosses represent 0.17 to 0.24 Na/Ca and 23 to 27°F approach to adiabatic saturation; squares represent 23 to 27°F without sodium additive.

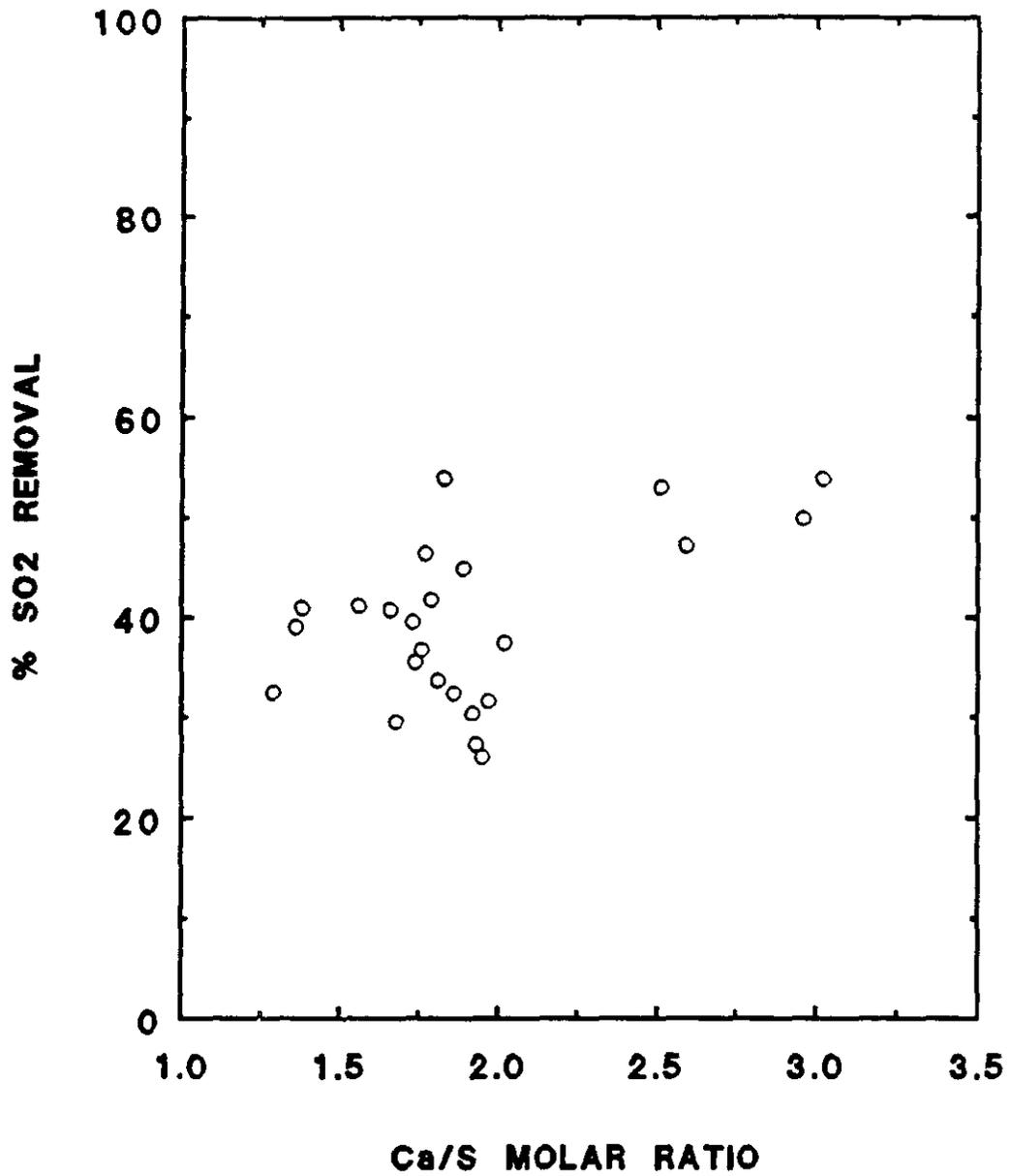


Figure 2. SO<sub>2</sub> removal as a function of Ca/S ratio for early tests using hydrated lime without sodium additive.

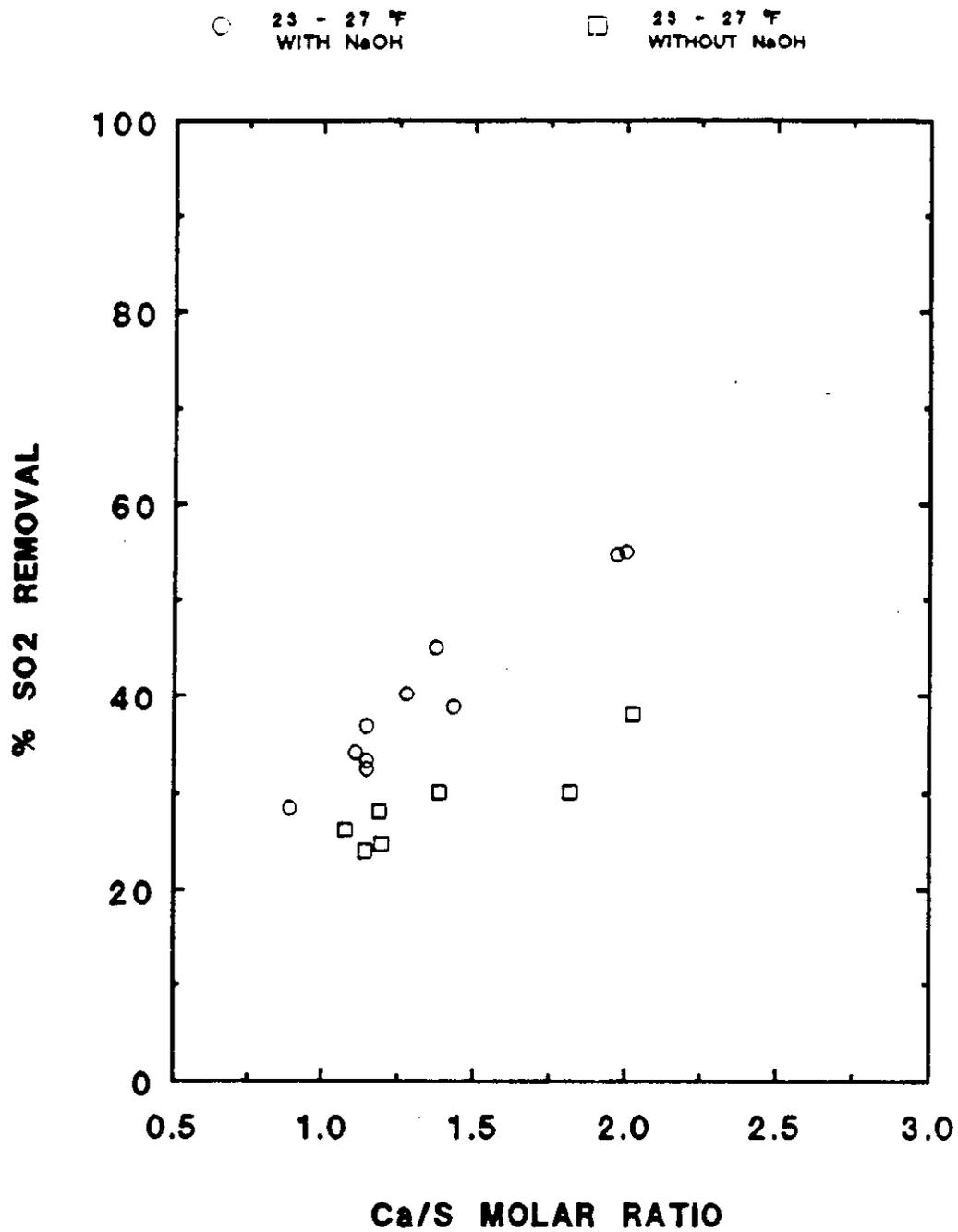


Figure 3. SO<sub>2</sub> removals observed using hydrated lime G at 23 to 27°F approach to adiabatic saturation temperature. Circles represent 0.17 to 0.24 Na/Ca; squares represent tests without sodium additive.

The SO<sub>2</sub> removals were calculated from the SO<sub>2</sub> concentrations measured at the humidifier inlet and ESP outlet using continuous gas analyzers which were corrected to dry, excess-air-free conditions. Corrections for air in-leakage were made using continuous oxygen analyzer data collected at both locations. The moisture content was calculated based on measured wet bulb and dry bulb temperatures. The Ca/S ratio was calculated based on the measured SO<sub>2</sub> concentration in the flue gas entering the humidifier, the measured flue gas flow rate into the humidifier, and the measured hydrated lime feed rate to the humidifier. More detailed information on the data measurement and analysis methods is provided in the "Demonstration Test Methods" subsection.

The sorbent utilization decreased somewhat with increasing Ca/S ratio. Sorbent utilization is calculated using the following formula:

$$\% \text{ Sorbent Utilization} = \frac{\% \text{ SO}_2 \text{ Removal}}{\text{Ca/S} (1 + 0.5 \text{ Na/Ca})} \quad (1)$$

where Ca/S and Na/Ca are molar ratios. Figure 4 shows the sorbent utilizations for the data shown in Figure 1 for hydrated lime A, plotted against the Ca/S ratio. At 19 to 22°F approach, the sorbent utilization averaged 33% at 1.4 Ca/S and 31% at 2.0 Ca/S. At 23 to 26°F, the sorbent utilizations were lower and averaged 29% at 1.4 Ca/S and 26% at 2.2 Ca/S. Figure 5 shows the sorbent utilizations for hydrated lime G plotted against the Ca/S mol ratio. The utilization was 26% at 1.1 Ca/S and 24% at 2.0 Ca/S which indicates a slightly lower sorbent utilization at the higher Ca/S ratio. The observed dependence of the SO<sub>2</sub> removal and the sorbent utilization on the Ca/S ratio was similar to that observed in 0.1 MW<sub>e</sub> pilot plant tests.<sup>1</sup> Comparison of pilot plant and Edgewater results is detailed later in this section.

#### Effect of Sodium Hydroxide Additive

The sodium additive increased both the SO<sub>2</sub> removal and the sorbent utilization. The increase in the SO<sub>2</sub> removal results from two causes. One is the co-sorbent effect of NaOH capturing SO<sub>2</sub>. The other is the promotional effect of sodium compounds to increase SO<sub>2</sub> capture by hydrated lime. The co-sorbent

effect can be estimated from the stoichiometric ratio in the reaction of  $\text{SO}_2$  and  $\text{NaOH}$  to form  $\text{Na}_2\text{SO}_3$  and  $\text{Na}_2\text{SO}_4$ .

The average  $\text{SO}_2$  removals for tests with and without sodium additive for hydrated limes A and G at constant Ca/S and approach to adiabatic saturation temperature are compared in Table 1. As the data in Table 1 show, sorbent utilization (Equation 1) increased with addition of the sodium additive. The values in Table 1 for tests without additive represent limited data (one to three runs at each condition), because the project goal was to demonstrate maximum  $\text{SO}_2$  removal. Additionally, with lime G most tests without  $\text{NaOH}$  were performed because the  $\text{NaOH}$  feed system was not available due to operating problems and, thus, were not performed in a planned block of tests. Table 1 shows the promotional effect of sodium for the tests using hydrated lime A. For example, at 2.0 Ca/S and 19 to 22°F approach,  $\text{SO}_2$  removal averaged 70% with sodium additive, but only 44% without sodium additive (a difference of 26% (absolute)). This was 15% higher than incremental  $\text{SO}_2$  removal by the co-sorbent effect (115% stoichiometric utilization, Table 1). These results clearly indicate the promotional effect of the added sodium on  $\text{SO}_2$  removals using lime A.

The promotional effect of sodium with hydrated lime G was not observed as indicated in Table 1 by 83 to 85% stoichiometric additive utilization for hydrated lime G. The promotional effect of sodium additive on  $\text{SO}_2$  removal performance was established in pilot plant studies with both hydrated limes.<sup>2-4</sup> The full-scale results using hydrated lime A at Edgewater confirm the additive promotional effect. Based on the pilot test data, the additive promotional effect using lime G was expected to be similar, although the demonstration test data did not confirm the promotional effect.

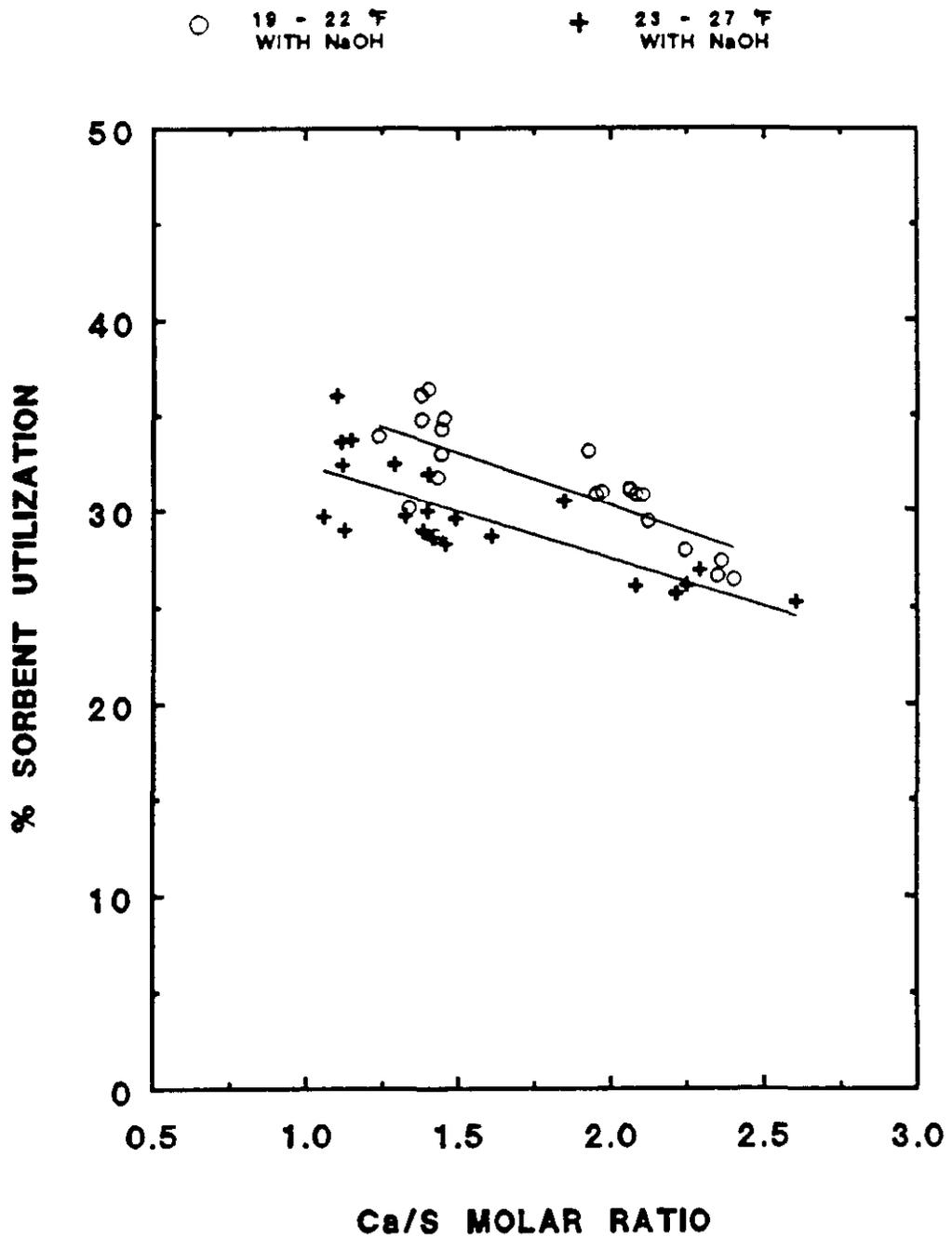


Figure 4. Sorbent utilization versus Ca/S ratio using hydrated lime A at 0.17 to 0.24 Na/Ca molar ratio. Circles represent 19 to 22°F approach to adiabatic saturation temperature; crosses represent 23 to 27°F approach.

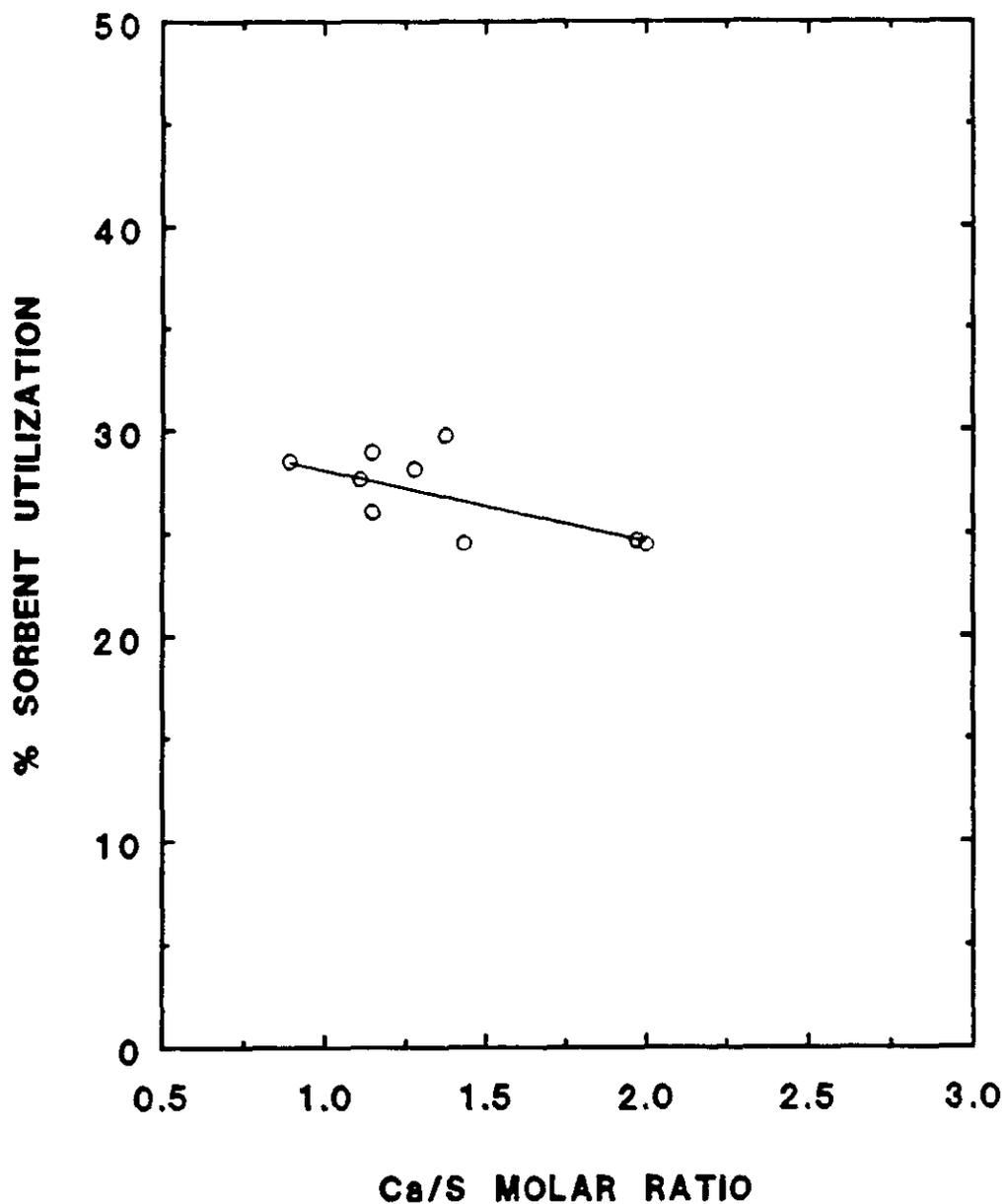


Figure 5. Sorbent utilization versus Ca/S ratio using hydrated lime G at 23 to 27°F approach to adiabatic saturation temperature. The Na/Ca molar ratio was 0.17 to 0.24.

**TABLE 1**  
**EFFECT OF NaOH ADDITION**

Hydrated Lime	Conditions		Desulfurization Performance				
	Ca/S, mol/mol	Approach, °F	Avg % SO <sub>2</sub> Removal		Percent Stoichio- metric Additive Utiliza- tion (c)	Average % Sorbent Utilization (a)	
			Without Additive	With Additive (b)		Without Additive	With Additive (b)
A	1.8-2.2	19-22	44	70	115	22	31
A	1.8-2.2	23-27	35	62	118	18	27
A	1.0-1.2	23-27	30	42	107	27	34
G	1.8-2.0	23-27	34	55	83	18	25
G	1.3-1.4	23-27	29	41	85	21	27

(a) Sorbent Utilization = 
$$\frac{\% \text{ SO}_2 \text{ Removal}}{\text{Ca/S} \left( 1 + 0.5 \frac{\text{Na}}{\text{Ca}} \right)}$$

(b) 0.18 to 0.25 Na/Ca molar ratio

(c) Percent Stoichiometric Additive Utilization = 
$$\frac{\% \text{ Removal with Additive} - \% \text{ Removal Without Additive}}{0.5 \text{ Na/Ca}}$$

### Effect of Approach to Adiabatic Saturation Temperature

At constant Ca/S and Na/Ca ratios, SO<sub>2</sub> removal was higher when the process was operated at closer approach to adiabatic saturation (or wet bulb) temperature. The effect of only a few degrees variation in the approach to adiabatic saturation on SO<sub>2</sub> removal can be observed by comparing the circles (19 to 22°F approach) with the crosses (23 to 27°F approach) for hydrated lime A in Figure 1. This comparison shows that, at equivalent Ca/S ratios, the observed SO<sub>2</sub> removals were 6 to 10 percentage points (absolute) higher in the tests at closer approach conditions. The effect of a wider variation in the approach to saturation is shown in Figure 6. Although some variation occurred in the observed SO<sub>2</sub> removals at similar approach temperatures, the data demonstrate that the SO<sub>2</sub> removal increases with decreasing approach to adiabatic saturation temperature. At 1.8 to 2.2 Ca/S, the SO<sub>2</sub> removal was about 70% at 20°F approach, but only 60% at 30°F approach. At 1.2 to 1.6 Ca/S, the SO<sub>2</sub> removal was about 50% at 20°F approach, but only about 35% at 30°F approach. Figure 7 shows a similar effect of approach to adiabatic saturation on SO<sub>2</sub> removal in tests using hydrated lime G.

In the tests without additive made early in the test program (August 28 to September 28, 1989) using hydrated lime A, the desired approach to adiabatic saturation temperature was 20°F. However, based on the variation in the flue gas temperature continuously measured at the inlet of the ESP, the actual approach may have varied between 20°F to 30°F. To best observe the effect of approach to adiabatic saturation on SO<sub>2</sub> removal for these early tests, the removals are plotted against the ESP inlet temperature in Figure 8. The data cover a Ca/S molar ratio range of 1.7 to 2.0. Despite some scatter, Figure 8 shows a trend to higher removals at lower ESP inlet temperatures (closer approach to adiabatic saturation temperature). The observed SO<sub>2</sub> removal increased from 25% to nearly 55% as the ESP inlet temperature was decreased from 152°F to 142°F. This range of temperature variation was roughly equivalent to a variation in the approach to adiabatic saturation from 20°F to 30°F. The humidifier inlet gas analyzers were not operational during these early tests. Therefore, the SO<sub>2</sub> removals were calculated using the stack gas analyzer baseline data on flue gas SO<sub>2</sub> content taken before and after the tests. This method should be accurate for

determining the SO<sub>2</sub> removal efficiency as long as there are not large variations in the coal sulfur content or lime feed during the test. The ESP inlet temperature is used because the humidifier outlet temperature was sometimes affected by wet deposits that would build up on the outlet thermocouples and their shields. This caused uncertainty in the humidifier outlet temperature and, thus, the approach estimation. When the deposition was light, the thermocouples at the humidifier outlet and ESP inlet agreed to within 1 to 2°F. When the deposition was heavier, the ESP inlet thermocouples responded faster than the humidifier outlet thermocouples to process changes. This sometimes resulted in larger observed temperature differences. During subsequent testing, the deposition problem was reduced by better process control. This is covered in more detail in Section 4.

Variations in the approach to adiabatic saturation were not intended as part of the demonstration test program. The variations shown in Figures 6 through 8 occurred for two reasons. First, the approach varied because of variations in the humidifier exit temperature from the control point. The five shielded thermocouples used for measuring the humidifier exit gas temperature were positioned in front of the exit turning vanes; initially, these were the humidifier exit temperature control thermocouples. After about three months of operation, however, five additional unshielded thermocouples were installed downstream of the turning vanes and humidifier exit temperature control was transferred to these "turning vane" thermocouples. This change was made to make process control less sensitive to deposit formation on the humidifier exit thermocouples. The humidifier exit thermocouples agreed with the turning vane thermocouples to within 3°F. This level of precision is considered normal for temperature control instrumentation using uncalibrated thermocouples. This margin of agreement, however, means that the actual humidifier outlet temperature could vary from the desired set point by up to 3°F. Combined with the typical reliability limit in the wet bulb reading ( $\pm 2^\circ\text{F}$ ), the approach variation could become even wider. The second reason for the variations shown in Figures 6 and 7 is that the set point for the approach to adiabatic saturation was increased from 20°F to 25°F during the tests with hydrated lime A. This change was necessary because of humidifier performance deterioration as discussed in

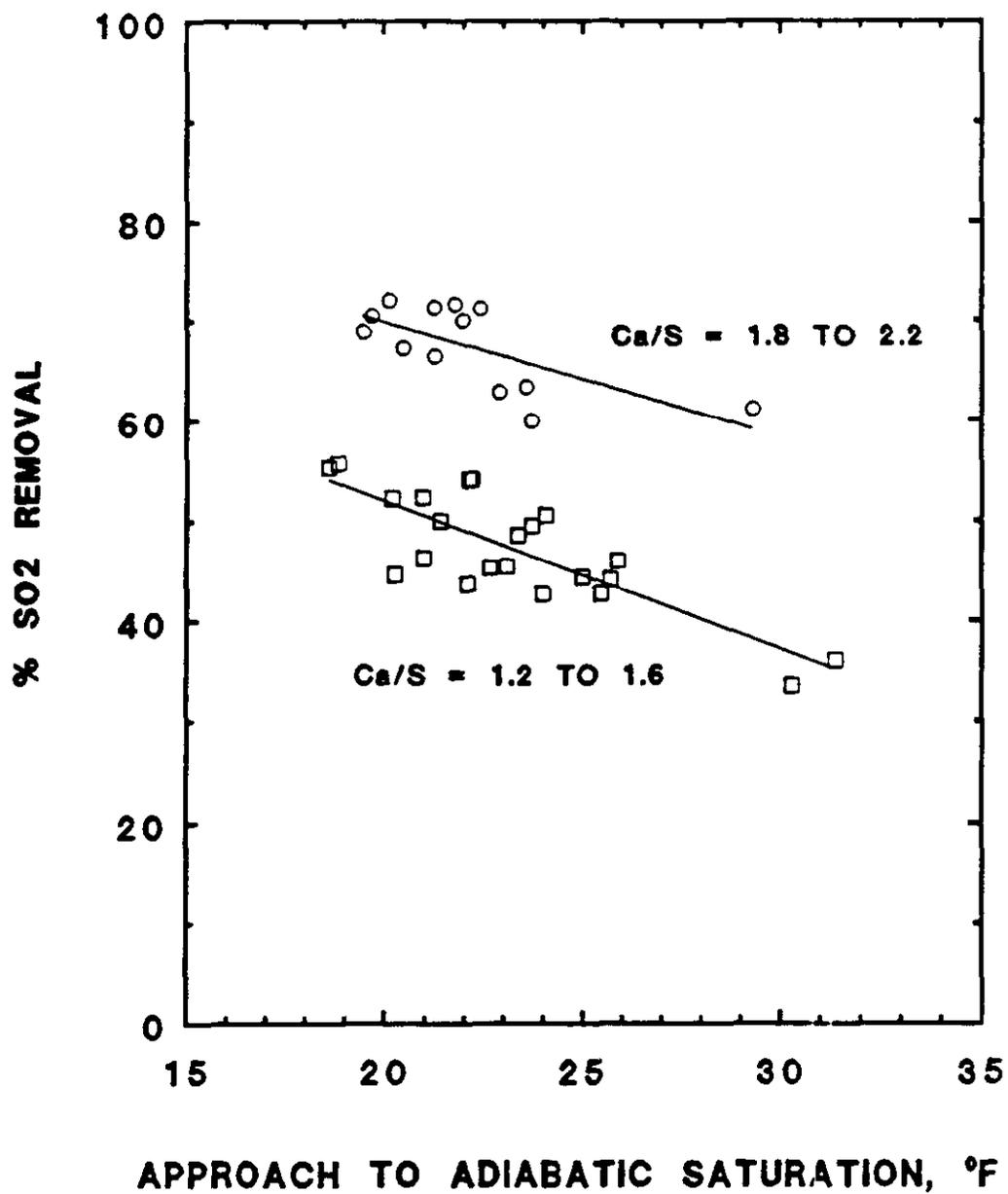


Figure 6. SO<sub>2</sub> removal versus approach to adiabatic saturation temperature for tests using hydrated lime A. The Na/Ca mol ratio was 0.17 to 0.24.

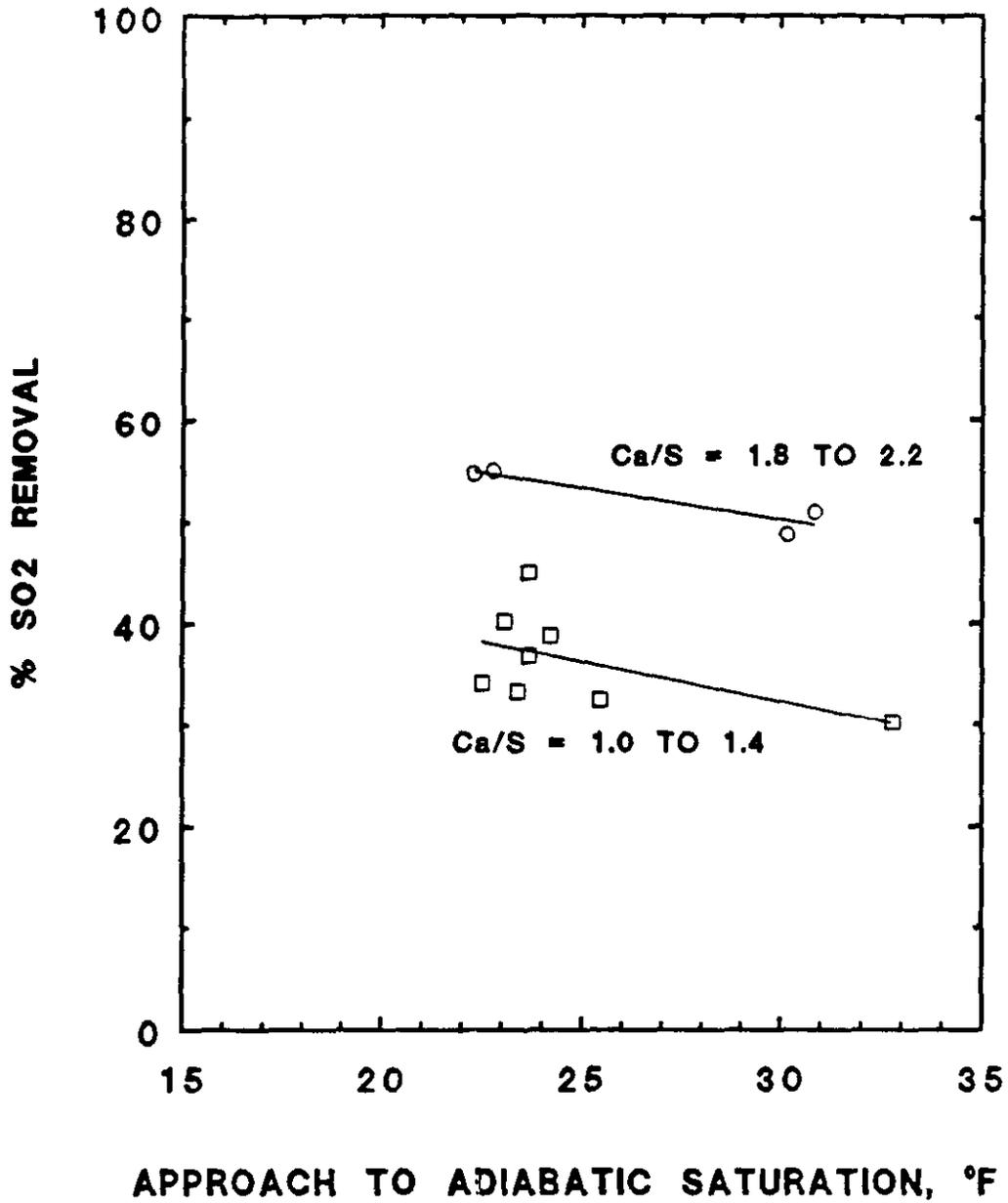


Figure 7. SO<sub>2</sub> removal versus approach to adiabatic saturation temperature for tests using hydrated lime G. The Na/Ca mol ratio was 0.17 to 0.24.

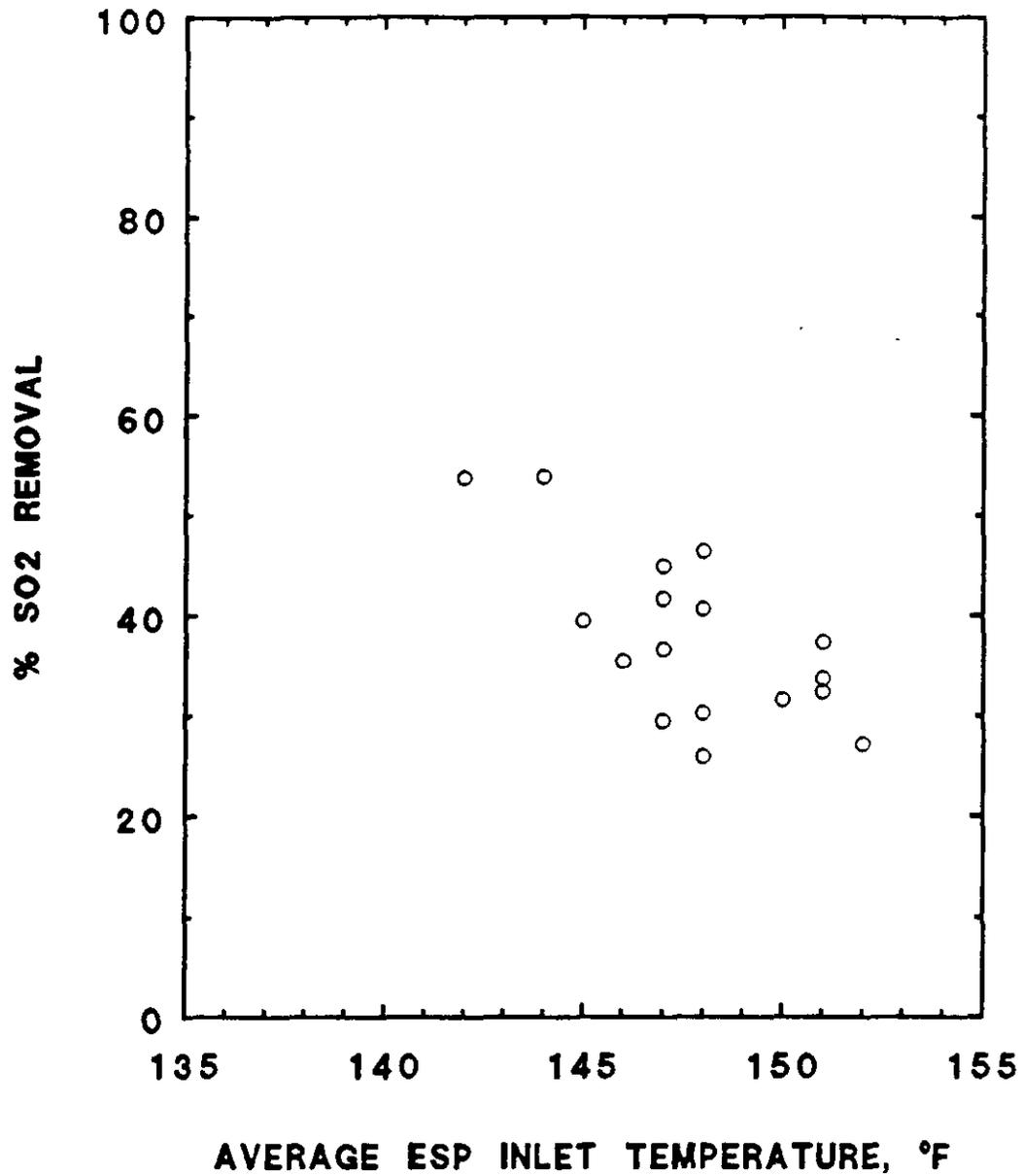


Figure 8. SO<sub>2</sub> removal versus average ESP inlet temperature for early tests using hydrated lime A without sodium additive.

Section 4. The tests using hydrated lime G were all made at 25°F approach to the adiabatic saturation set point.

#### Effect of Different Hydrated Limes

The choice of hydrated lime can affect the level of SO<sub>2</sub> removal at similar process conditions, as was shown in pilot plant tests.<sup>1,5</sup> Typical lime analyses are given in Table 2. Both are high calcium (>88% Ca(OH)<sub>2</sub> by weight) hydrated limes. Differences in physical properties, such as surface area, may contribute to the performance differences. The surface areas were 22 to 24 m<sup>2</sup>/g for hydrated lime A and 15 to 18 m<sup>2</sup>/g for hydrated lime G. Previously reported work showed a correlation between sorbent surface area and SO<sub>2</sub> removal performance.<sup>4</sup> A complete list of hydrated lime analyses, including surface area and thermogravimetric analyses (TGA), is given in Appendix A. Figure 9 shows the observed SO<sub>2</sub> removals plotted against Ca/S ratio for tests using hydrated lime A (crosses) and hydrated lime G (squares); common conditions were 0.17 to 0.24 Na/Ca mol ratio and 23 to 27°F approach to adiabatic saturation temperature. Comparison of the crosses with the squares show that at equivalent Ca/S ratios, the observed SO<sub>2</sub> removals were 5 to 10 percentage points (absolute) higher when using hydrated lime A than when using hydrated lime G. These results are consistent with pilot plant results that showed higher SO<sub>2</sub> removals when using hydrated lime A than when using hydrated lime G.<sup>1</sup>

The reactivity difference between the two hydrated limes was not as apparent in tests without NaOH additives. At 2.0 Ca/S and 23 to 27°F approach, the SO<sub>2</sub> removals were 35% using hydrated lime A and 34% using hydrated lime G. At 1.4 Ca/S the SO<sub>2</sub> removals were 32% using lime A and 29% using lime G.

Ten runs were made during shakedown testing using a third hydrated lime (lime H) without sodium addition. However, because this was a shakedown test period, not enough runs were made at any single condition to make worthwhile comparisons.

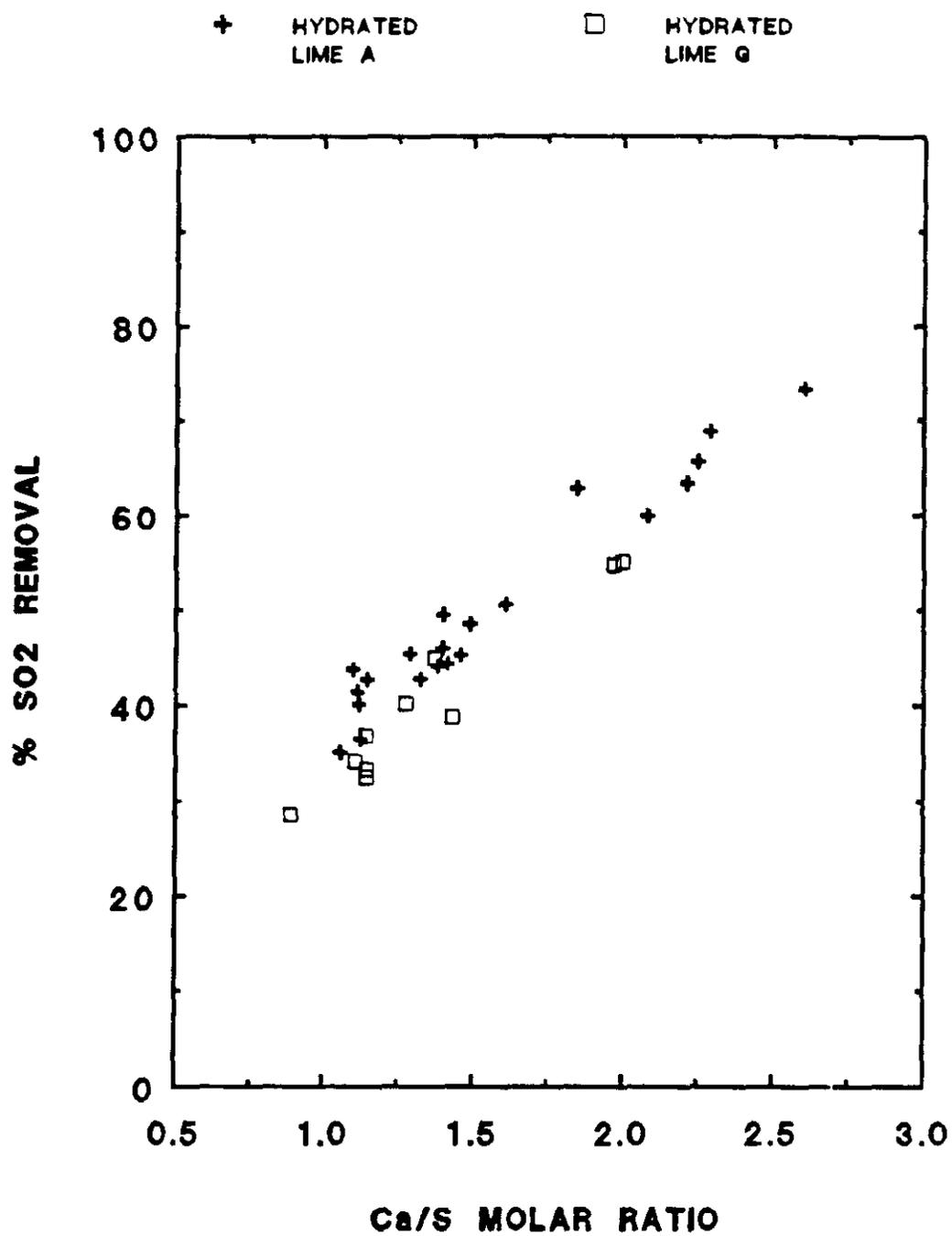


Figure 9. Comparison of SO<sub>2</sub> removals using hydrated limes A and G. The common conditions were 0.17 to 0.24 Na/Ca and 23 to 27°F approach to adiabatic saturation temperature.

## Other Variables and Statistical Analysis

In addition to the three primary process variables (Ca/S, Na/Ca, approach to adiabatic saturation temperature) discussed in the previous sections, the humidifier inlet gas SO<sub>2</sub> concentration and the humidifier inlet gas temperature were identified during pilot plant tests as secondary process variables that affect SO<sub>2</sub> removal. However, the Edgewater data indicate that the observed SO<sub>2</sub> removals were not sensitive to either variable. It is not clear if this insensitivity is real, or the result of test data variability obscuring the small secondary variable effects. No other process variables had a significant effect on SO<sub>2</sub> removal.

Analysis of the effect of other possible variables was based on the variables shown to be significant in pilot plant tests. The pilot plant results showed that two secondary process variables, humidifier inlet flue gas temperature and SO<sub>2</sub> content, had statistically significant effects on SO<sub>2</sub> removal. The pilot plant inlet temperature test range was 270 to 330°F and the inlet SO<sub>2</sub> content test range was 540 to 2700 ppmv (dry). Removal of SO<sub>2</sub> increased with increasing flue gas inlet temperature. This was likely due to the increased quantity of water spray required to humidify the hotter flue gas. The increased water spraying possibly increased droplet sorbent interactions, which are known to enhance SO<sub>2</sub> capture in the humidifier.<sup>6</sup> The pilot plant SO<sub>2</sub> removal decreased with increasing SO<sub>2</sub> content in the inlet flue gas, which may be related to the decreasing ratio of water droplets to sorbent particles with increasing SO<sub>2</sub> content at the same Ca/S ratio. With a lower droplet-to-particle ratio, a smaller fraction of sorbent particles may interact with droplets. This would reduce SO<sub>2</sub> removal in the humidifier based on previous studies.<sup>6</sup>

The test program at Edgewater was not designed with the intention of measuring the effect of these two secondary variables on SO<sub>2</sub> removal. However, variations in both variables occurred during Edgewater testing. In once-through tests without recycle sorbent, the humidifier inlet gas temperature varied between 260 and 290°F, depending on the boiler operating conditions, the boiler soot blowing schedule and the air preheater efficiency. The humidifier inlet gas SO<sub>2</sub> concentration varied between 700 and 1850 ppmv (dry), depending on the coal

**TABLE 2**  
**TYPICAL HYDRATED LIME ANALYSES**

Hydrated Lime	Mois- ture wt %	Ash, dry wt %	Ash Elementals, dry wt %						Lime Index %	TGA Data		BET* Surface Area m <sup>2</sup> /g
			SiO <sub>2</sub>	Al <sub>2</sub> O <sub>3</sub>	Fe <sub>2</sub> O <sub>3</sub>	CaO	MgO	Na <sub>2</sub> O		Ca(OH) <sub>2</sub> , wt %	CaCO <sub>3</sub>	
A	0.3	75.83	1.31	0.20	0.11	94.22	0.90	0.01	92.8	92.97	2.48	23.2
G	0.7	76.83	2.07	0.55	0.30	93.44	3.03	0.03	93.0	87.95	2.48	16.7

\*Brunauer, Emmett, and Teller, J. Am. Chem. Soc., 60, 309 (1938)

sulfur content and the amount of excess air in the flue gas. A statistical analysis of the Edgewater data did not confirm the effect observed in the pilot plant tests of these secondary variables. A description of the statistical analysis follows.

A standard statistical t-test was performed on the data for once-through tests without recycle sorbent injection in the following manner. A least-squares linear regression equation was developed for SO<sub>2</sub> removal using the primary and secondary process variables listed in Table 3. Each variable's regression coefficient was divided by the coefficient's standard error to obtain the t-statistic for each variable. The t-statistic was used to test the significance of the process variables on SO<sub>2</sub> removal (t-test). A higher absolute value of t indicates a stronger dependence of SO<sub>2</sub> removal on the given process variable. The significance of each variable effect was determined by comparing the calculated t-values with tabulated values of Student's t distribution for the corresponding degrees of freedom.

The t-statistics and variable significance confidence levels are listed in Table 3. The results show the expected significant effect of the three primary variables (Ca/S, Na/Ca and approach) on SO<sub>2</sub> removal for both hydrated limes A and G, in agreement with pilot plant results. The results concerning the two secondary variables did not confirm the pilot plant results.

The regression equation for hydrated lime A is:

$$\begin{aligned} \% \text{ SO}_2 \text{ removal} = & 5.352 + 0.0004 X_1 + 0.0515 X_2 + 20.34 X_3 \\ & + 72.41 X_4 - 0.581 X_5 \end{aligned} \quad (2)$$

The regression equation for hydrated lime G is:

$$\begin{aligned} \% \text{ SO}_2 \text{ removal} = & 106.34 + 0.0022 X_1 - 0.370 X_2 + 15.97 X_3 \\ & + 53.75 X_4 - 0.165 X_5 \end{aligned} \quad (3)$$

where:

X<sub>1</sub> = Inlet SO<sub>2</sub> concentration, ppmv

$X_2$  = Inlet gas temperature, °F  
 $X_3$  = Ca/S, mol/mol  
 $X_4$  = Na/Ca, mol/mol  
 $X_5$  = Approach to adiabatic saturation temperature, °F

These regression equations are specific to these two data sets, which were obtained from the Edgewater demonstration tests which were not statistically designed.

Table 3 shows the coefficient of determination ( $R^2$ ) for the regression equations. This coefficient measures how well the regression equations fit the data. The data indicate that the variables listed in the table account for 92% ( $R^2 = 0.917$ ) of the variation in the  $SO_2$  removals using lime A and 89% ( $R^2 = 0.892$ ) of the variation in the  $SO_2$  removals using lime G. These relatively high coefficients indicate that the effects of any other variables would not have accounted for much of the variation in the observed  $SO_2$  removals.

The effect on  $SO_2$  removal of the inlet gas  $SO_2$  concentration (and, thus, the coal sulfur content) was not significant for either hydrated lime in the statistical analysis. This indicates that the level of  $SO_2$  removal was not sensitive to the coal sulfur content at Edgewater. As an illustration, the  $SO_2$  removals for tests using hydrated lime A at 1.0 and 2.0 Ca/S are plotted against the inlet gas  $SO_2$  concentration in Figure 10 and against the coal sulfur content in Figure 11; the common conditions were 0.17 to 0.24 Na/Ca and 19 to 26°F approach to adiabatic saturation. The trend lines in both figures are horizontal, indicating no correlation.

The effect of the humidifier inlet gas temperature also did not agree with the pilot plant results. In tests using hydrated lime A, the effect of inlet gas temperature was not significant in the statistical analysis. In tests using hydrated lime G, the effect appeared to be significant, with an opposite effect from that expected based on pilot plant tests. The negative t-value in Table 3

TABLE 3  
 STATISTICAL ANALYSIS OF PROCESS VARIABLE EFFECTS  
 ON SO<sub>2</sub> REMOVAL

Variable	Hydrated Lime A		Hydrated Lime G	
	t-Value* for SO <sub>2</sub> Removal	Variable** Signifi- cance, %	t-Value* for SO <sub>2</sub> Removal	Variable** Signifi- cance, %
No. of Runs	66		25	
Degrees of Freedom	60		19	
R <sup>2</sup>	0.917		0.892	
Inlet SO <sub>2</sub> concentration, dry ppmv	-0.13	NS	0.10	NS
Inlet gas temperature, °F	0.74	NS	-2.74	97.5-99
Ca/S in Humidifier, mol/mol	11.37	>99.5	6.95	>99.5
Na/Ca, mol/mol	6.86	>99.5	7.31	>99.5
Approach to adiabatic saturation, °F	-5.612	>99.5	-0.97	75-90

NS = Not significant to at least 75% confidence

\*t-Value =  $b/S_b$  where

b = coefficient in linear regression equation

S<sub>b</sub> = standard error of b

\*\*Percentage confidence that the variable has a significant effect on SO<sub>2</sub> removal, based on comparison with tabulated values of Student's t distribution at the corresponding degrees of freedom.

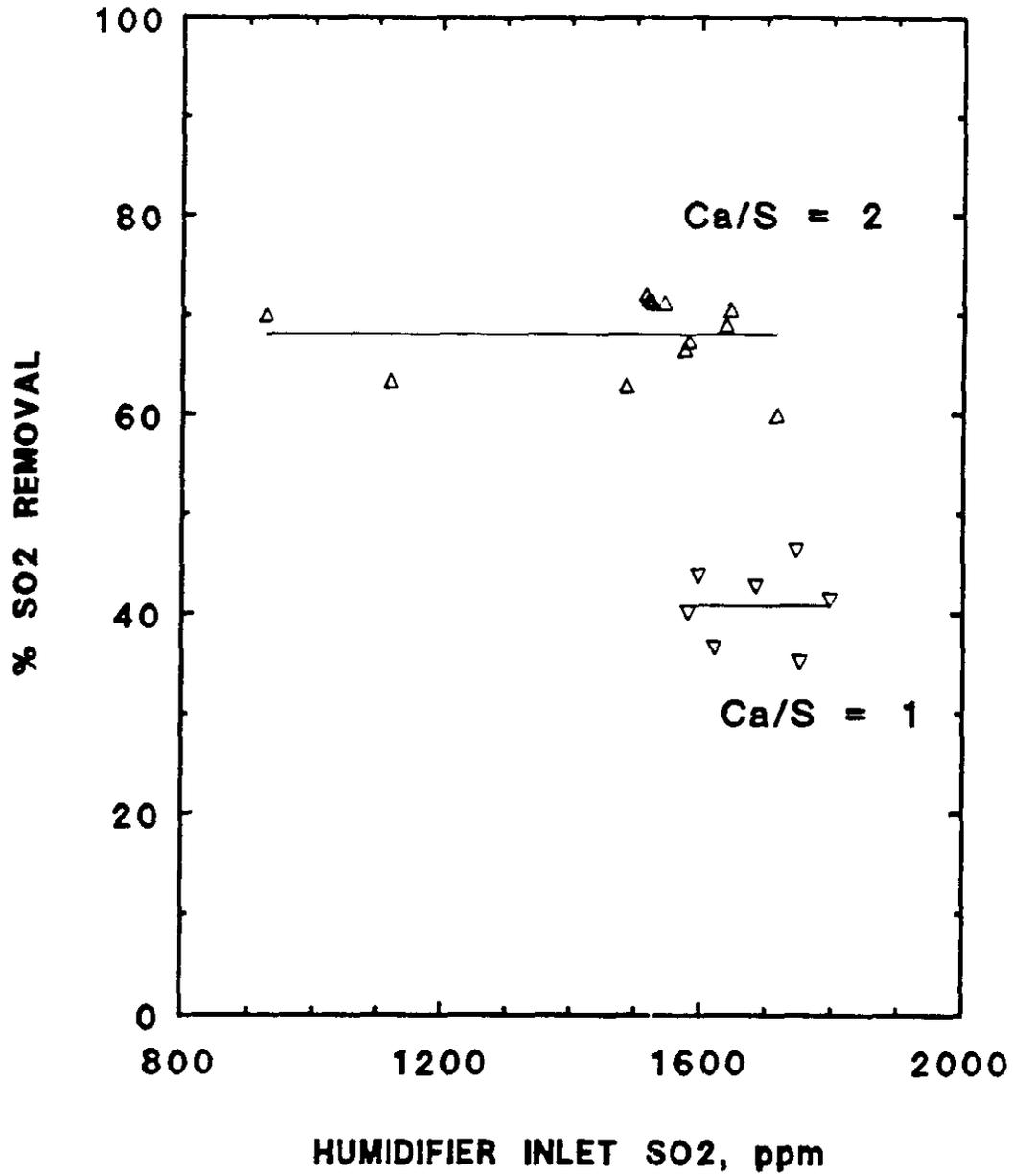


Figure 10. SO<sub>2</sub> removal versus SO<sub>2</sub> concentration in the flue gas at the humidifier inlet. Common conditions: hydrated lime A, 0.17 to 0.24 Na/Ca mol ratio, 19 to 27°F approach to adiabatic saturation temperature.

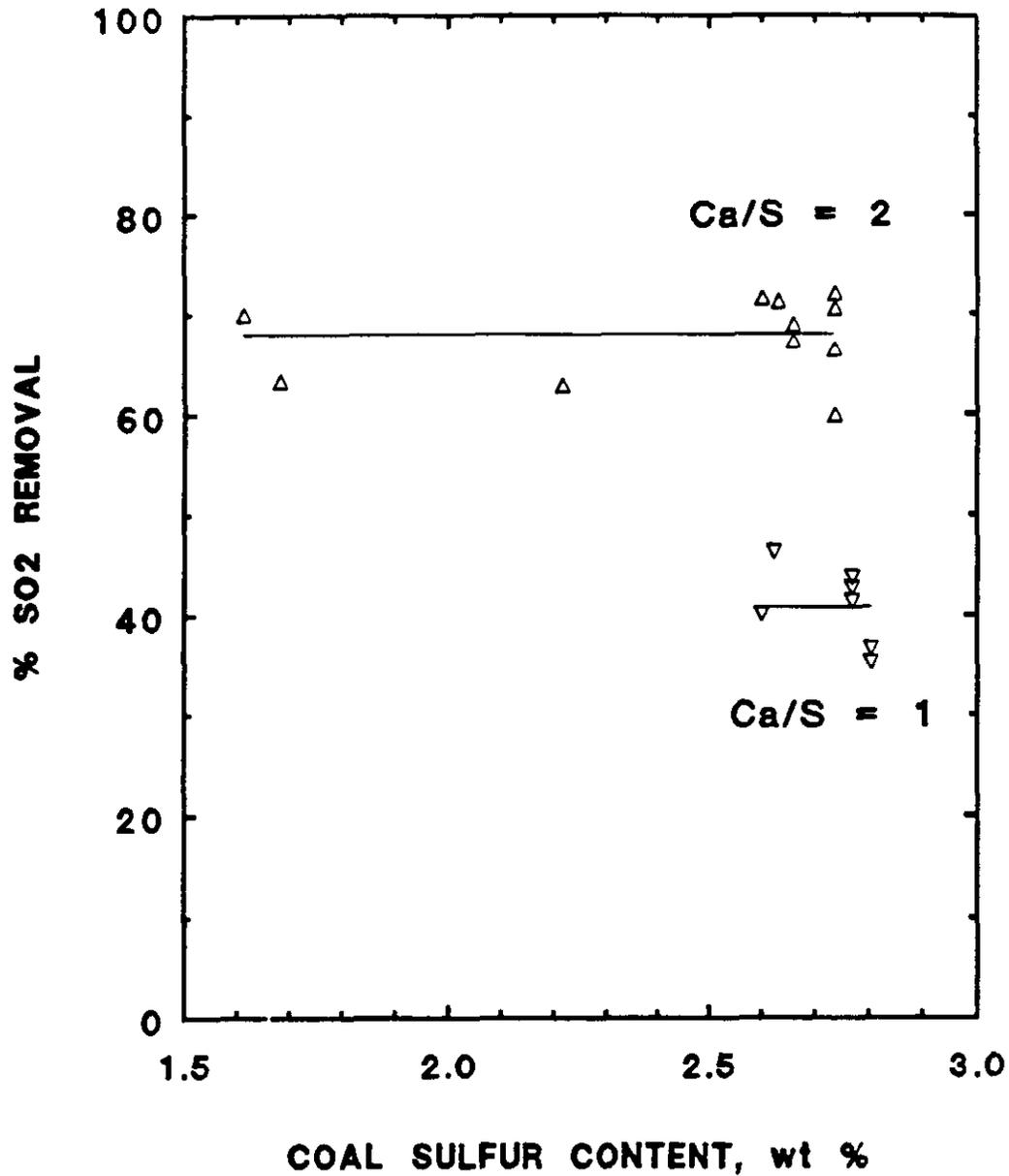


Figure 11. SO<sub>2</sub> removal versus coal sulfur content using hydrated lime A, at 19 to 27°F approach to adiabatic saturation temperature, with 0.17 to 0.24 Na/Ca mol ratio.

for inlet temperature using lime G would suggest that the effect of inlet temperature on SO<sub>2</sub> removal was negative, i.e., as the temperature increased, the SO<sub>2</sub> removal decreased. However, the inlet temperature was significantly correlated, or "coupled", with the approach to saturation in tests using hydrated lime G. In other words, the tests at higher inlet gas temperature were usually performed at higher approaches to adiabatic saturation. Thus, the apparent negative effect of inlet temperature was actually the effect of the coupling of this variable with the approach to saturation. Therefore, the effect of the humidifier inlet gas temperature on lime G cannot be determined because it cannot be separated from the effect of the approach to adiabatic saturation. Except for this one case, variable coupling did not occur for the other variables. The inlet gas temperature was not coupled with the approach, or any other variable, for the tests using hydrated lime A. Neither the inlet SO<sub>2</sub> concentration nor any of the primary variables were systematically coupled with any variable for either hydrated lime. The coupling of inlet temperature with the approach in the tests with hydrated lime G was coincidental, not the result of any specific operating procedure.

It is not clear if the lack of sensitivity to the secondary variables was due to the test data variability overshadowing the secondary effects, or if the lack of sensitivity was a real result of humidifier design differences. The differences between the pilot plant and the Edgewater unit include nozzle size and lime injector position differences. The water spray nozzles were larger, and they were in a 10 x 10 array at Edgewater, compared to a single nozzle in the center of the pilot unit humidifier. Thus, the spray patterns were different, as well as the droplet sizes. The lime injector position (relative to the nozzle position) was different for the two units, which probably resulted in different particle-droplet interactions. The effects of these differences on SO<sub>2</sub> removal are not clear.

#### SO<sub>2</sub> Removals During Bypass Operation

When some of the flue gas bypassed the humidifier, the observed system (humidifier plus ESP) SO<sub>2</sub> removals were lower than the SO<sub>2</sub> removals observed at similar process conditions during non-bypass operation. This was due to 1) a bypass effect in which a portion of the flue gas is not Coolside-treated, and

2) a reduction of the ESP's contribution to SO<sub>2</sub> removal because of the increased ESP gas temperature. By factoring out the bypass effect, the ESP's contribution to the total SO<sub>2</sub> removal during non-bypass operation was estimated. Table 4 shows the calculated humidifier SO<sub>2</sub> removals for tests during bypass operation. Observed system SO<sub>2</sub> removals from tests at similar conditions without bypass are shown for comparison. In each case, the calculated humidifier SO<sub>2</sub> removals were lower than the observed system SO<sub>2</sub> removals from tests without bypass at similar conditions. This indicates that 5 to 15% (absolute) of the total system SO<sub>2</sub> removal may occur in the ESP when flue gas does not bypass the humidifier. The data indicate that most of the SO<sub>2</sub> removal occurs in the humidification zone.

During bypass operation, SO<sub>2</sub> removals were lower than during tests in which all of the flue gas passed through the humidification chamber at similar humidity conditions. For example, at 2.0 Ca/S and 0.2 Na/Ca and 23 to 27°F approach using hydrated lime A, the SO<sub>2</sub> removal was 60 to 65% without bypass. However, when 35% of the flue gas bypassed the humidifier, the total system SO<sub>2</sub> removal was 36% at 2.0 Ca/S, based on the flue gas flow in the humidifier. This lower SO<sub>2</sub> removal during bypass was the result of two factors. First, bypassed flue gases were not treated by the Coolside process to remove SO<sub>2</sub>. Second, the SO<sub>2</sub> removal contributed by the ESP is reduced since approach to adiabatic saturation in the ESP is increased as a result of combining the hot, non-humidified bypassed gas with the humidifier exit gas before the ESP. If the ESP SO<sub>2</sub> capture is assumed to be negligible because of the increased flue gas temperature, the humidifier SO<sub>2</sub> removals can be calculated using the following equation:

$$\Delta SO_2 (\text{hum}) = \frac{\Delta SO_2 (\text{sys})}{1 - \frac{\%BP}{100}} \quad (4)$$

where:

$\Delta SO_2 (\text{hum})$  is the calculated SO<sub>2</sub> removal in the humidifier

$\Delta SO_2 (\text{sys})$  is the measured total system SO<sub>2</sub> removal

%BP is the percentage of flue gas which bypassed the humidifier, based on the average gas temperatures measured at the humidifier inlet and outlet and ESP inlet

TABLE 4  
BYPASS OPERATION RESULTS

<u>Hydrated Lime</u>	<u>Ca/S, Based on Flue Gas Flow in Humidifier</u>	<u>Na/Ca, mol</u>	<u>Approach to Adiabatic Saturation, °F</u>	<u>Percent Bypass</u>	<u>% SO<sub>2</sub> Removal Total System (Measured)</u>	<u>Humidifier Only (Calculated)</u>
A	2.0	0.21	19-22	0	70	-
	2.0	0.23	19-22	25	38	50
A	2.0	0.22	23-27	0	62	-
	2.0	0.23	23-27	35	36	55
A	1.4	0.20	23-27	0	46	-
	1.5	0.20	23-27	28	28	39
A	2.0	0	23-27	0	35	-
	2.1	0	23-27	25	20	27
A	1.1	0	23-27	0	30	-
	1.2	0	23-27	20	12	14
G	2.0	0.25	23-27	0	55	-
	2.0	0.26	23-27	35	20	31
G	1.1	0.22	23-27	0	34	-
	1.1	0.25	23-27	36	10	16
G	1.9	0	23-27	0	34	-
	2.1	0	23-27	35	13	21
G	1.2	0	23-27	0	26	-
	1.0	0	23-27	28	8	11

The negligible ESP SO<sub>2</sub> capture is a reasonable assumption at high (25 to 35%) bypass rates, since the approach to adiabatic saturation in the ESP is about 60 to 70°F at these bypass conditions.

#### SO<sub>2</sub> REMOVAL RESULTS WITH SORBENT RECYCLE

Recycle sorbent showed a significant capacity for additional SO<sub>2</sub> capture. This was observed in a test involving injection of recycle sorbent only and in recycle process tests involving injection of both fresh and recycle sorbents. In the recycle process tests, the observed SO<sub>2</sub> removals were somewhat lower than in once-through sorbent tests at the same total available Ca/S ratios, indicating that the recycle sorbent was not as reactive as the fresh sorbent. Additional tests are needed to better evaluate the recycle effects, since the operating circumstances at Edgewater during the recycle process tests prevented the attainment of steady-state recycle conditions.

Most of the recycle process tests were performed using sorbent G at 24 to 30°F approach to adiabatic saturation. The recycle process tests were performed using a fresh hydrated lime feed rate that provided 0.7 to 0.8 fresh Ca/S molar ratio in the humidifier. The recycle sorbent provided an available Ca/S of 0.5 to 0.7 in the form of unreacted Ca(OH)<sub>2</sub>. The SO<sub>2</sub> removals for the recycle and no-recycle tests are shown in Figure 12. Since the no-recycle tests were made at slightly different approaches to saturation temperature, the observed SO<sub>2</sub> removal data also were affected by the difference in the approach to saturation. Also, the no-recycle tests were not performed below 1.0 Ca/S, so the no-recycle data must be extrapolated to 0.7 to 0.8 Ca/S for comparison. Extrapolation of the no-recycle data to 0.7 to 0.8 Ca/S gives an average SO<sub>2</sub> removal of about 22% at 23 to 27°F approach. The SO<sub>2</sub> removal using 8000 lb/hr recycle sorbent (0.5 to 0.7 available Ca/S) and 0.7 to 0.8 fresh Ca/S at 24 to 30°F approach ranged between 23 and 31%, with an average of 26%. Thus, the recycle sorbent did provide a significant additional SO<sub>2</sub> capture compared to once-through sorbent alone, even if the average 26% SO<sub>2</sub> removal in the recycle tests is directly compared without adjustment for the slightly higher approach used in the recycle tests.

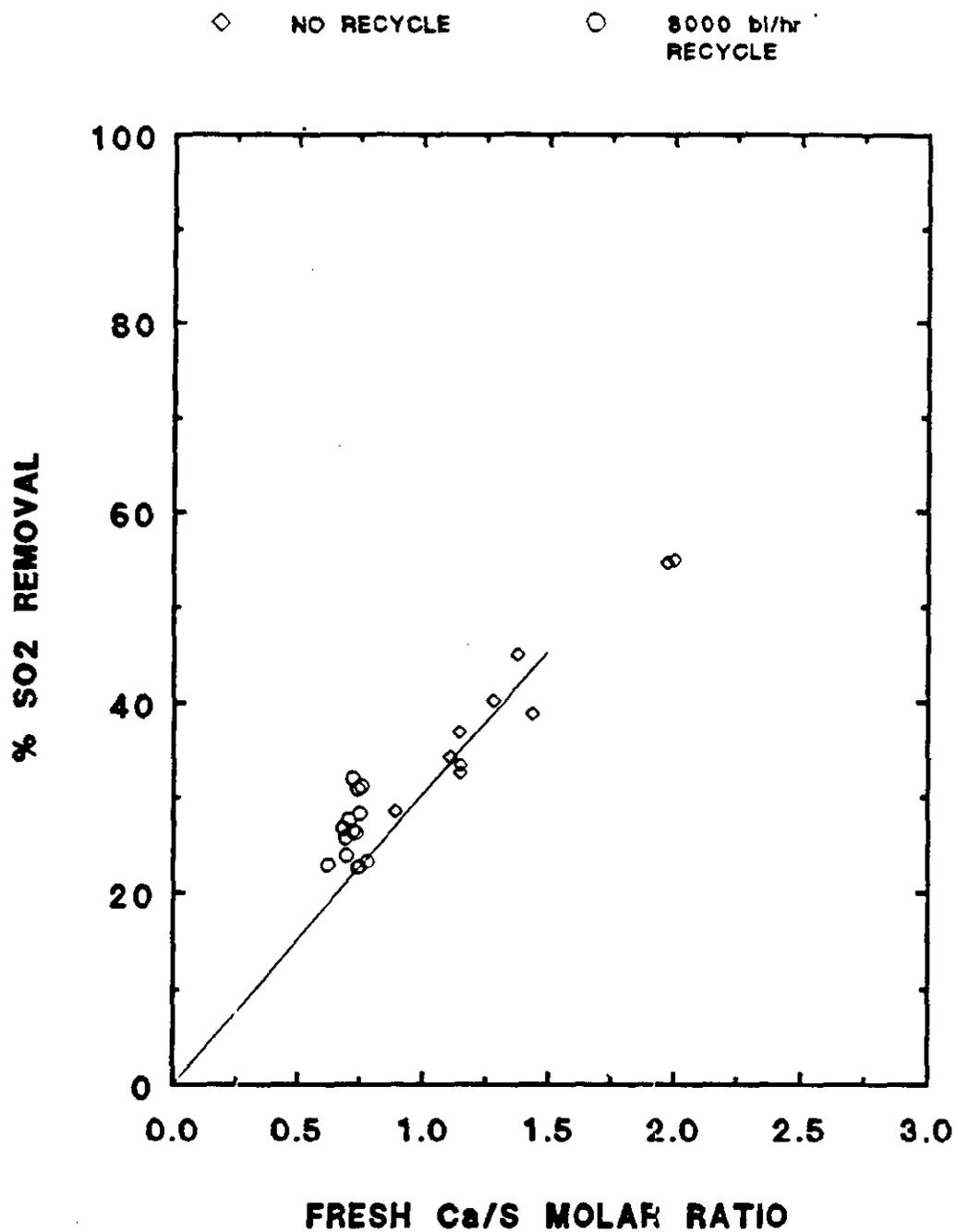


Figure 12. Comparison of SO<sub>2</sub> removals during once-through and recycle tests using hydrated lime G, 23 to 27°F approach, 0.17 to 0.24 fresh Na/Ca mol ratio. The solid line represents extrapolation of once-through data to low Ca/S ratio.

When considered on the basis of equivalent total available Ca/S (instead of equivalent fresh Ca/S), the SO<sub>2</sub> removal observed during recycle tests was less than that observed during once-through tests. The total available Ca/S ratio for the recycle tests was 1.2 to 1.5 (0.7 to 0.8 fresh Ca/S + 0.5 to 0.7 available Ca/S in the recycle sorbent). The available Ca in the recycle feed is equal to the unreacted Ca(OH)<sub>2</sub>. The 26% SO<sub>2</sub> removal observed in these tests was lower than the 39 to 45% SO<sub>2</sub> removal observed using only fresh hydrated lime at 1.3 to 1.5 Ca/S. The difference in the SO<sub>2</sub> removal could be due to the recycle sorbent being less reactive than the fresh sorbent on an equivalent Ca/S basis, or to the slightly higher approach used in the recycle tests.

Other recycle tests were performed using 0.8 fresh Ca/S and 8000 lb/hr recycle without fresh NaOH addition. The recycle sorbent contained significant sodium (0.10 to 0.20 Na/Ca by mol), however, and there was no basis for comparison with once-through tests.

One sorbent recycle test was performed in which no fresh hydrated lime (lime G) was fed to determine the level of SO<sub>2</sub> removal using recycle feed alone. The approach to adiabatic saturation was 18°F and the recycle sorbent feed rate was 7000 lb/hr. The recycle sorbent, which contained 10.7 wt % available Ca(OH)<sub>2</sub> and 0.20 Na/Ca mol ratio in the recycle solids, provided 0.5 available Ca/S. The observed SO<sub>2</sub> removal was 22%. This test indicates that the recycled sorbent alone removed a significant amount of SO<sub>2</sub> at 18°F approach. This demonstrated that recycling spent sorbent increases sorbent utilization.

Some recycle tests using hydrated lime A were performed, but the data were too limited for meaningful comparison.

During the recycle process tests, steady-state recycle conditions were not attained at Edgewater for two reasons. First, because flue gas bypass operation was necessary at high load conditions, the available calcium content of the ESP ash varied with boiler load. This resulted in lower sorbent utilization and in higher coal fly ash-to-sorbent ratios during bypass operations. Since the load usually changed two to four times daily, there never was a continuous production of uniform spent sorbent for longer than 12 to 16 hours. Second, when the

recycle tests were performed, the sodium hydroxide additive pump was out of service for two days. This caused variations in the sodium content of the recycle ash. The recycle ash analyses (Appendix A) show that the available calcium content varied between 10 and 20 wt %  $\text{Ca}(\text{OH})_2$  and the sodium content varied between 0.8 and 2.8 wt %  $\text{Na}_2\text{O}$ .

The recycle tests were the last set of tests performed. The performance deterioration of the water spray nozzles during the course of the demonstration program did not allow consistent operation of the recycle process tests at 20°F approach, as described elsewhere in this report ("Humidifier Chamber: Water Spray Lance Design and Operation" in Section 4). Although sorbent recycle shows promise, additional tests under better controlled conditions would be required to better evaluate the effect of sorbent recycle.

## DATA RELIABILITY AND COMPARISON WITH PILOT PLANT DATA

### Data Reliability

The ESP outlet gases were sampled using EPA-certified continuous flue gas analyzers. The humidifier inlet flue gas analyzer system accuracy was established using EPA Method 6. With no hydrated lime or water feed, the two analyzer systems agreed within a few percent. The data from these two continuous analysis systems were used to calculate the  $\text{SO}_2$  removals for most of the test program. Sorbent utilizations based on the ESP hopper ash sample analyses agreed with those calculated from the gas analyzer data and sorbent feed rate for the entire set of tests using hydrated lime A and for the recycle tests using hydrated lime G. The agreement for once-through tests using hydrated lime G was not as good. This may have been due to difficulty in attaining steady-state operation, resulting from frequent flue gas bypass and other minor operating problems, as discussed in Section 4.

The  $\text{SO}_2$  removals were based on the continuous gas analysis system readings on gases withdrawn from the flue gas using probes placed at the humidifier inlet and ESP exit. Each gas analyzer system consists of  $\text{SO}_2$  and  $\text{O}_2$  analyzers. These analyzers were calibrated once daily using standard gas and automatically spanned

and zeroed every four hours. The SO<sub>2</sub> and O<sub>2</sub> analyzers at the ESP exit were EPA-certified instruments for continuous analysis of stack gases for the power plant emissions measurements. A Du Pont SO<sub>2</sub> analyzer was used at the ESP exit for the first two months of the test program. Beginning September 27, 1989, a Western SO<sub>2</sub> analyzer was used. The O<sub>2</sub> analyzer at the ESP exit was manufactured by Thermo. Reliability of humidifier inlet SO<sub>2</sub> analyzer (ThermoElectron) and O<sub>2</sub> analyzer (Westinghouse) readings was verified using EPA Method 6 for SO<sub>2</sub> analysis. Five Method 6 tests were performed on October 12, 1989, over a 3.75-hour period, and the flue gas SO<sub>2</sub> measurements were compared with gas analyzer readings taken at the same time. A Relative Accuracy Test (RAT) indicated that the analyzers agreed with the manual sampling results. The RA is calculated as follows:

$$RA = \frac{\bar{d} + cc}{\overline{RM}} \times 100 \quad (5)$$

where:

RA is the relative accuracy in %

$\bar{d}$  is the absolute value of the mean differences between pair data sets

cc is the absolute value of the confidence coefficient (calculated from the standard deviation of the differences and set at 97.5% confidence)

$\overline{RM}$  is the average SO<sub>2</sub> value determined using EPA Method 6

Assuming typical test variability ( $\pm 7\%$ ), the relative accuracy for five tests must be less than 15.7%, for the two methods to be considered in agreement. Table 5 lists the test results and the RAT results. Although the results show a slight, but consistent bias (with the gas analyzer data showing an average of 69 ppm less SO<sub>2</sub> than the EPA method), the relative accuracy at Edgewater was 6%, which was well below the 15.7% RA required to show agreement. The flue gas SO<sub>2</sub> concentration, based on ESP exit analyzer readings, agreed with those based on the humidifier inlet analyzer readings to within 5% when lime and water were shut off, indicating good agreement between the two continuous analyzer systems.

A probe for withdrawing gas samples was installed at the humidifier exit. However, reactive solids in the humidified flue gas collected on this probe and

TABLE 5  
COMPARISON OF INLET SO<sub>2</sub> ANALYZER  
AND EPA METHOD 6 RESULTS

<u>Time</u>	<u>SO<sub>2</sub>, ppm</u>		<u>Difference</u>
	<u>EPA Method 6</u>	<u>Gas Analyzer Data</u>	
14:16 to 14:51	1568	1538	30
14:56 to 15:31	1697	1610	87
15:40 to 16:15	1679	1623	56
16:45 to 17:20	1763	1687	76
17:28 to 18:03	1749	1651	98
Averages	1691.2	1621.8	69.4
Standard Deviations	77.2	55.4	26.9

$$cc = \frac{26.9 \times 2.776}{\sqrt{5}} = 33.4$$

$$RA = 100\% \times \frac{69.4 + 33.4}{1691.2} = 6.08\%$$

continued to remove SO<sub>2</sub> from the sample gas stream. Thus, the gas collected using this probe was not an accurate representation of the gas in the duct and, therefore, the humidifier exit gas probe was not used:

The chemical analyses of ESP hopper ash samples agreed with gas analyzer data for the entire set of tests using hydrated lime A and the recycle tests using hydrated lime G. Agreement was not as good for the once-through tests using hydrated lime G. Figures 13 to 15 compare sorbent utilizations determined from analyses of the ESP hopper ash samples with sorbent utilizations based on process run conditions (hydrated lime and NaOH feeds and measured SO<sub>2</sub> removal). Sorbent utilization based on process run conditions was described earlier (Equation 1). Sorbent utilization was determined for ESP hopper samples using the following equation:

$$\% \text{ Utilization} = \frac{\text{wt \% S}/32}{\text{wt \% CaO}/56 + \text{wt \% Na}_2\text{O}/62} \times 100 \quad (6)$$

where the S, CaO, and Na<sub>2</sub>O were determined on an as-received basis. Correction for the contribution of the coal ash to the CaO and Na<sub>2</sub>O contents were made based on their ratios to the SiO<sub>2</sub> and Al<sub>2</sub>O<sub>3</sub> contents in the coal ash. This is possible because the levels of SiO<sub>2</sub> and Al<sub>2</sub>O<sub>3</sub> in the feed hydrated lime are very small and so do not contribute much to the total solids. The average CaO/SiO<sub>2</sub> and CaO/Al<sub>2</sub>O<sub>3</sub> ratios in the coal ash were 0.029 and 0.061, respectively. The average Na<sub>2</sub>O/SiO<sub>2</sub> and Na<sub>2</sub>O/Al<sub>2</sub>O<sub>3</sub> ratios were 0.009 and 0.020, respectively. Multiplying the ESP hopper sample SiO<sub>2</sub> content by 0.029 gives the CaO from the coal ash, based on SiO<sub>2</sub> content; multiplying the ESP hopper sample Al<sub>2</sub>O<sub>3</sub> content by 0.061 gives the CaO from the coal ash, based on the Al<sub>2</sub>O<sub>3</sub> content. The average of the two values was subtracted from the ESP hopper ash CaO content. The Na<sub>2</sub>O content was corrected in a similar manner.

The parity plots in Figures 13 and 14 show good agreement between the two methods for the once-through tests using hydrated lime A and for the recycle tests using hydrated lime G. The agreement is not as good for once-through tests using hydrated lime G, as shown in Figure 15. Statistical analysis confirms

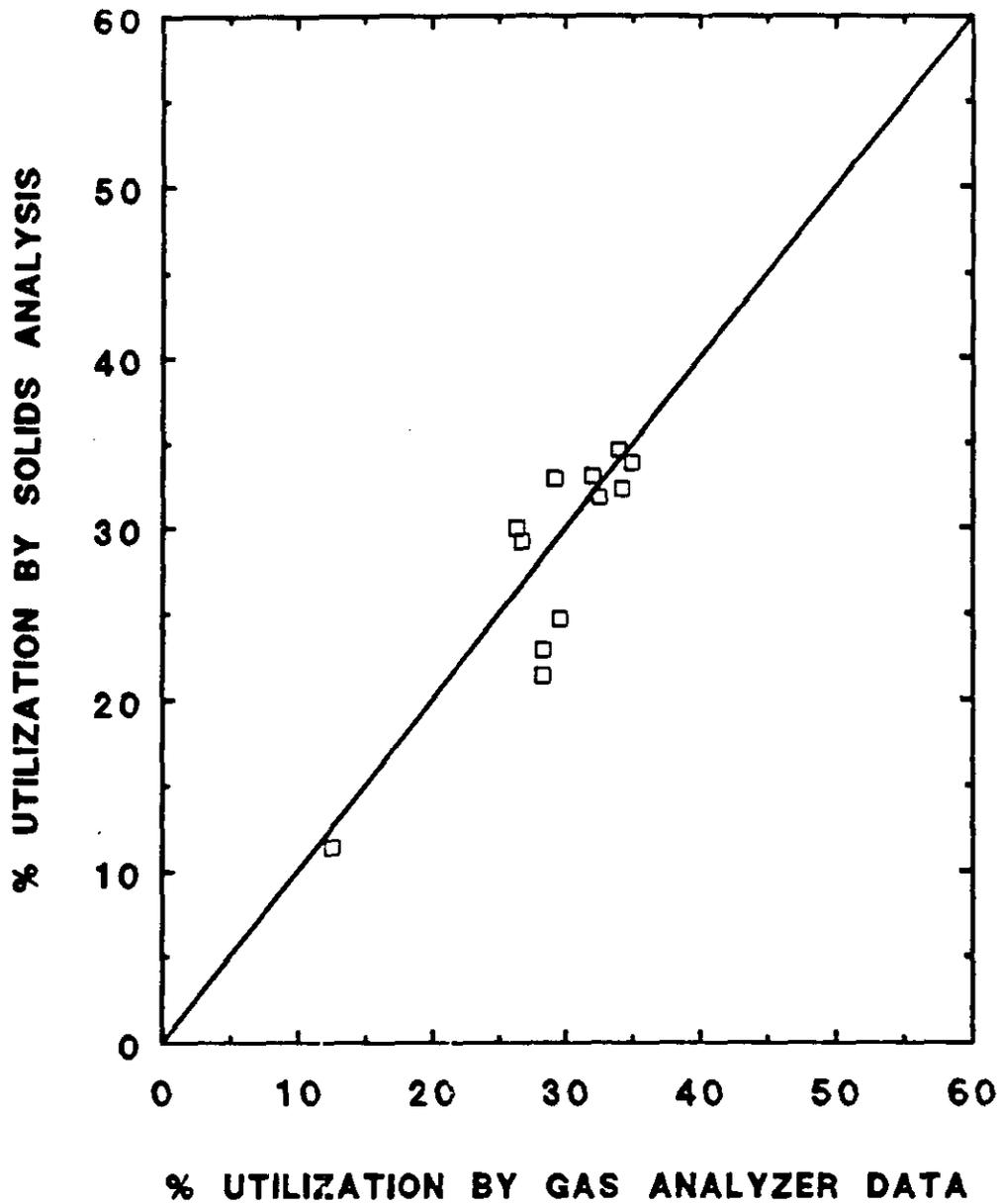


Figure 13. Comparison of sorbent utilization based on ESP hopper solids analysis with utilization based on gas analyzer data and lime and NaOH feed rates for hydrated lime A tests.

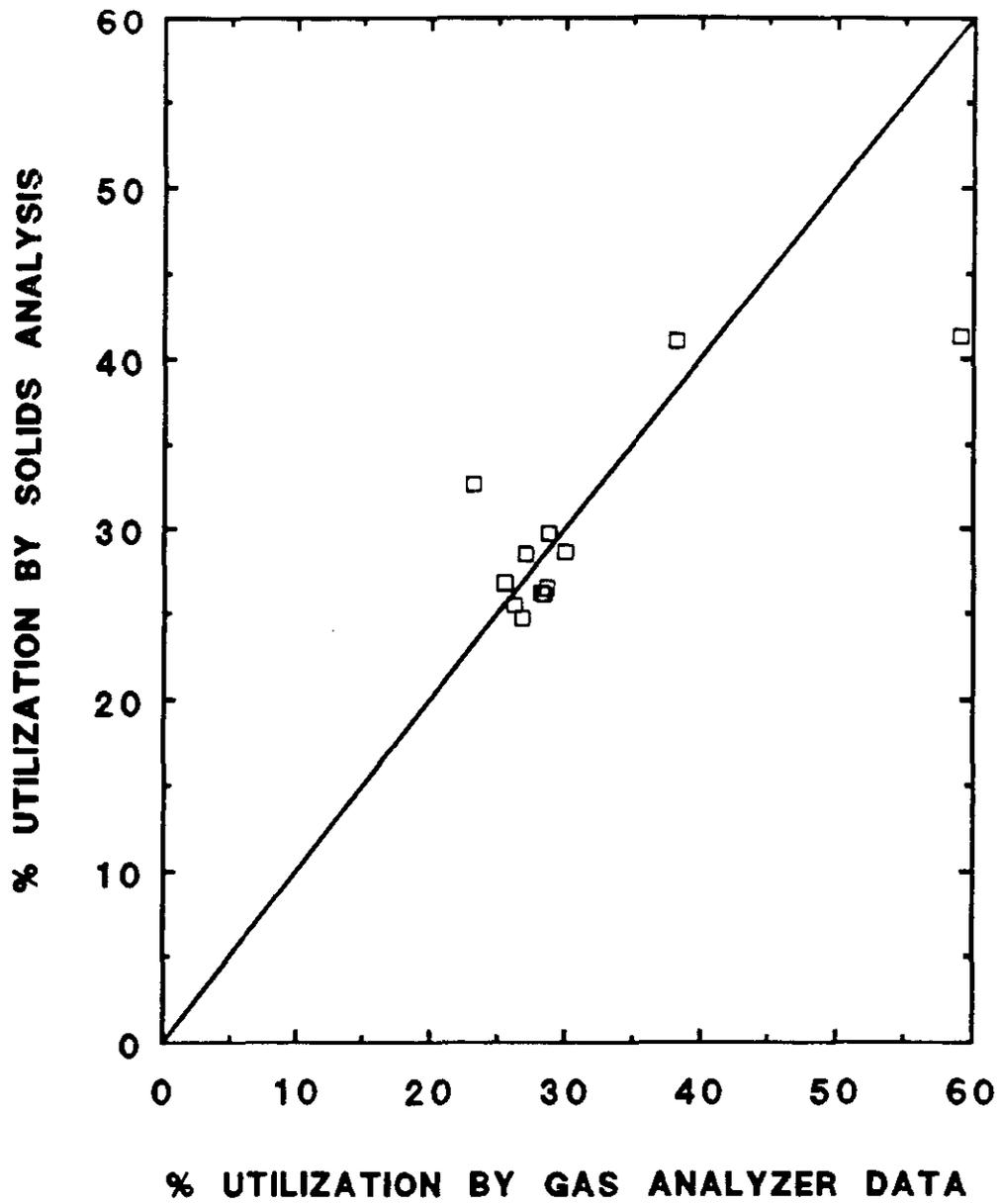


Figure 14. Sorbent utilization comparison for recycle tests using hydrated lime G.

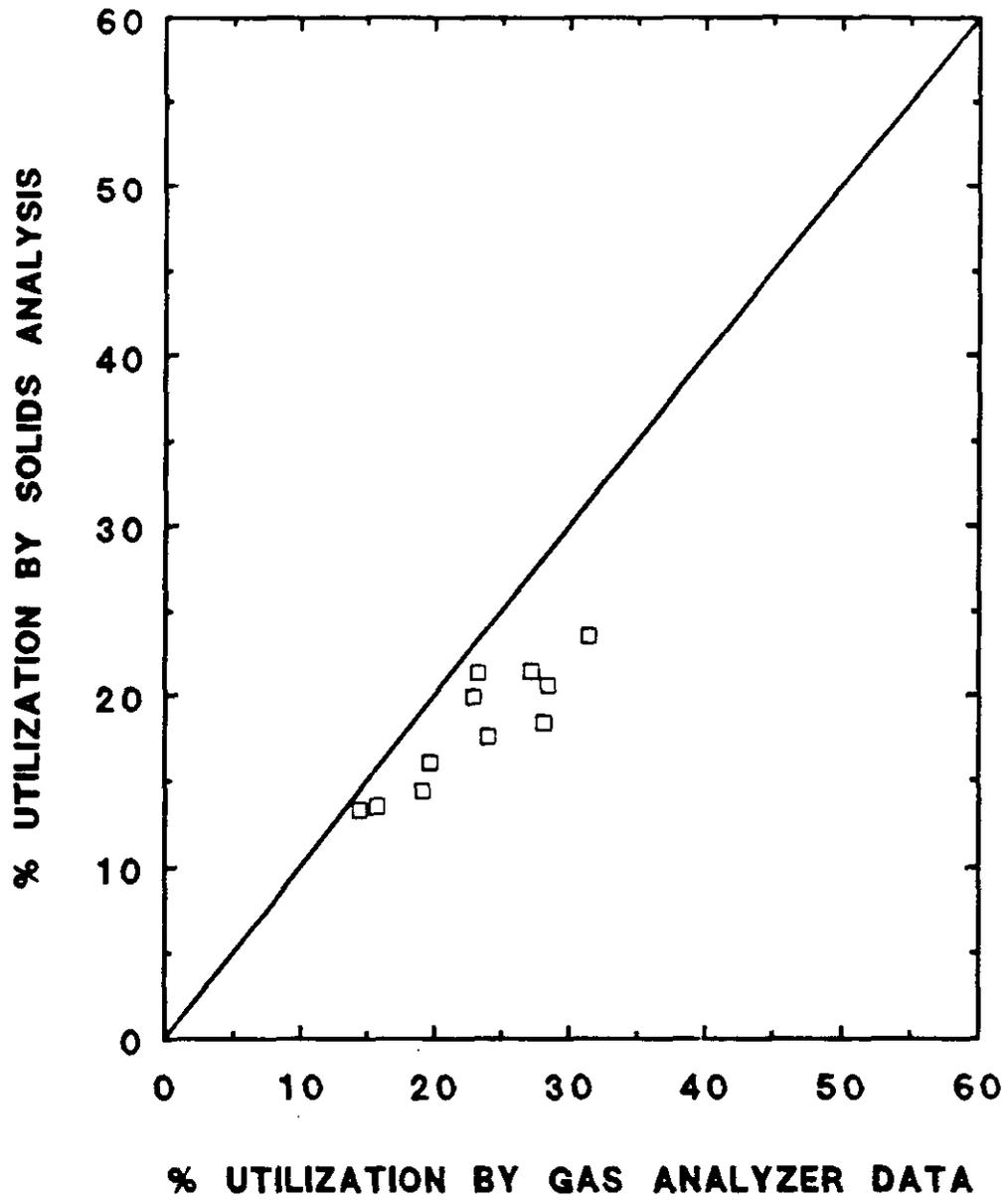


Figure 15. Sorbent utilization comparison for once-through tests using hydrated lime G.

**TABLE 6**  
**COMPARISON OF ESP ASH ANALYSES AND PROCESS RUN DATA**  
**HYDRATED LIME A, ONCE-THROUGH TESTS**

<u>Ca/S (mol)</u>	<u>Na/Ca (mol)</u>	<u>% SO<sub>2</sub> Removal</u>	<u>% Sorbent Utilization</u> <u>Based on</u>		<u>Difference</u>
			<u>SO<sub>2</sub> Removal,</u> <u>Ca/S and Na/Ca*</u>	<u>Ash</u> <u>Analysis**</u>	
1.56	0.00	41.1	26.3	30.0	-3.7
1.89	0.19	58.6	28.3	22.9	5.4
1.21	0.28	46.9	34.0	34.6	-0.6
1.29	0.17	44.7	31.9	31.8	0.1
1.45	0.18	52.7	33.3	32.3	1.0
1.45	0.18	53.7	34.0	33.8	0.2
1.40	0.21	48.2	31.2	33.0	-1.8
2.05	0.23	57.8	25.3	29.2	-3.9
1.49	0.11	45.7	29.1	24.7	4.4
1.96	0.23	60.8	27.8	32.9	-5.1
1.03	0.00	29.1	28.3	21.4	6.9
2.17	0.00	27.1	12.5	11.4	<u>1.1</u>
<b>Average</b>					<b>0.33</b>

\* 
$$\frac{\% \text{ SO}_2 \text{ Removal}}{\text{Ca/S} + 0.5 (\text{Na/Ca})}$$

\*\* 
$$\frac{\text{Total Sulfur}/32}{\text{CaO}/56 + \text{Na}_2\text{O}/62}$$
, CaO and Na<sub>2</sub>O corrected for calcium and sodium in coal ash

**TABLE 7**  
**COMPARISON OF ESP ASH ANALYSES AND PROCESS RUN DATA**  
**HYDRATED LIME G, RECYCLE TESTS**

Fresh Ca/S (mol)	Fresh Na/Ca (mol)	% Bypass	lb/hr. Recycle Feed	% SO <sub>2</sub> Removal	% Sorbent Utilization Based on		Difference
					SO <sub>2</sub> Removal, Ca/S and Na/Ca Recycle Feed and Bypass*	Ash Analysis**	
0.793	0.176	0	9108	28.3	26.8	24.7	2.1
0.799	0.181	0	9611	31.1	28.2	26.2	2.0
0.760	0.167	0	9159	32.0	30.0	28.6	1.4
0.721	0.163	0	7672	26.8	28.8	29.7	-0.9
0.761	0.163	0	9119	26.5	28.6	26.5	2.1
0.658	0.168	0	7609	22.9	28.4	26.1	2.3
1.126	0.0	0	8011	22.1	23.2	32.6	-9.4
0.785	0.183	0	8581	22.7	25.5	26.8	-1.3
0.814	0.187	0	8730	24.7	26.2	25.5	0.7
0.772	0.194	0	8280	26.3	27.1	28.5	-1.4
0.825	0.192	0	7361	39.7	38.2	41.0	-2.8
0.0	0.0	0	6827	21.7	59.1	41.3	<u>17.8</u>
Average							1.05

\* 
$$\frac{\% \text{ SO}_2 \text{ Removal}}{\text{Ca/S} + 0.5 (\text{Na/Ca})}$$

\*\* 
$$\frac{\text{Total Sulfur}/32}{\text{CaO}/56 + \text{Na}_2\text{O}/62}$$
, CaO and Na<sub>2</sub>O corrected for calcium and sodium in coal ash

**TABLE 8**  
**COMPARISON OF ESP ASH ANALYSES AND PROCESS RUN DATA**  
**HYDRATED LIME G, ONCE-THROUGH TESTS**

<u>Ca/S (mol)</u>	<u>Na/Ca (mol)</u>	<u>% SO<sub>2</sub> Removal</u>	<u>% Sorbent Utilization</u>		<u>Difference</u>
			<u>Based on</u>		
			<u>SO<sub>2</sub> Removal,</u> <u>Ca/S and Na/Ca*</u>	<u>Ash</u> <u>Analysis**</u>	
1.26	0	24.8	19.7	16.1	3.6
1.92	0	30.2	15.7	13.5	2.2
1.96	0	28.2	14.4	13.3	1.1
1.35	0.224	40.2	26.8	18.4	8.4
1.35	0.216	34.2	22.9	19.9	3.0
2.11	0.241	55.0	23.3	21.3	2.0
0.94	0.241	28.6	27.2	21.4	5.8
1.45	0.185	45.0	28.4	20.6	7.8
1.39	0.296	50.1	31.4	23.5	7.9
1.44	0.000	27.5	19.1	14.4	4.7
1.88	0.167	48.8	24.0	17.6	6.4
Average					4.81

\* 
$$\frac{\% \text{SO}_2 \text{ Removal}}{\text{Ca/S} + 0.5 (\text{Na/Ca})}$$

\*\* 
$$\frac{\text{Total Sulfur}/32}{\text{CaO}/56 + \text{Na}_2\text{O}/62}$$
, CaO and Na<sub>2</sub>O corrected for calcium and sodium in coal ash

this. Tables 6 to 8 list the sorbent utilizations based on the two methods. A standard statistical F test was performed using the method of variance over residual variances. For the once-through test data using hydrated lime A (Table 6), the differences between the results of the two methods of determining sorbent utilization were not significant. Using the 12 process runs and the two methods of determining utilization as sources of variance, the F-number for the method variance/residual variance was 0.67 for 1/11 degrees of freedom. This indicates that the probability that the two methods gave truly different results was insignificant. The results for the data from recycle tests using hydrated lime G (Table 7) were similar. Using the 12 process runs and the two methods of determining utilization as sources of variance, the F-number was 0.33 for 1/11 degrees of freedom, which indicates that the probability that the two methods gave truly different results was not significant for the recycle tests using hydrated lime G. The same F-test method performed on the once-through tests using hydrated lime G (Figure 15), however, gave different results. Using the 11 process runs shown in Table 8 and the two methods for determining utilization as sources of variance, the F-number was 31.5 for 1/10 degrees of freedom, which indicates that the results of the two methods were significantly different for this set of tests.

#### Comparison with Pilot Plant Data

System SO<sub>2</sub> removals were generally lower at Edgewater than in the Consol 0.1 MW<sub>e</sub> pilot plant under similar process conditions. The probable reason for the difference is the more efficient gas/solid contact in the pilot plant baghouse compared to that in the Edgewater ESP. Differences in other process design parameters may have contributed to the observed performance difference. For example, one of the parameters, important for SO<sub>2</sub> removal efficiency but not optimized in the Edgewater design, is the hydrated lime distribution in the flue gas.

In the pilot plant once-through simulation tests, least squares regression equations were developed for SO<sub>2</sub> removal in the humidifier and across the entire

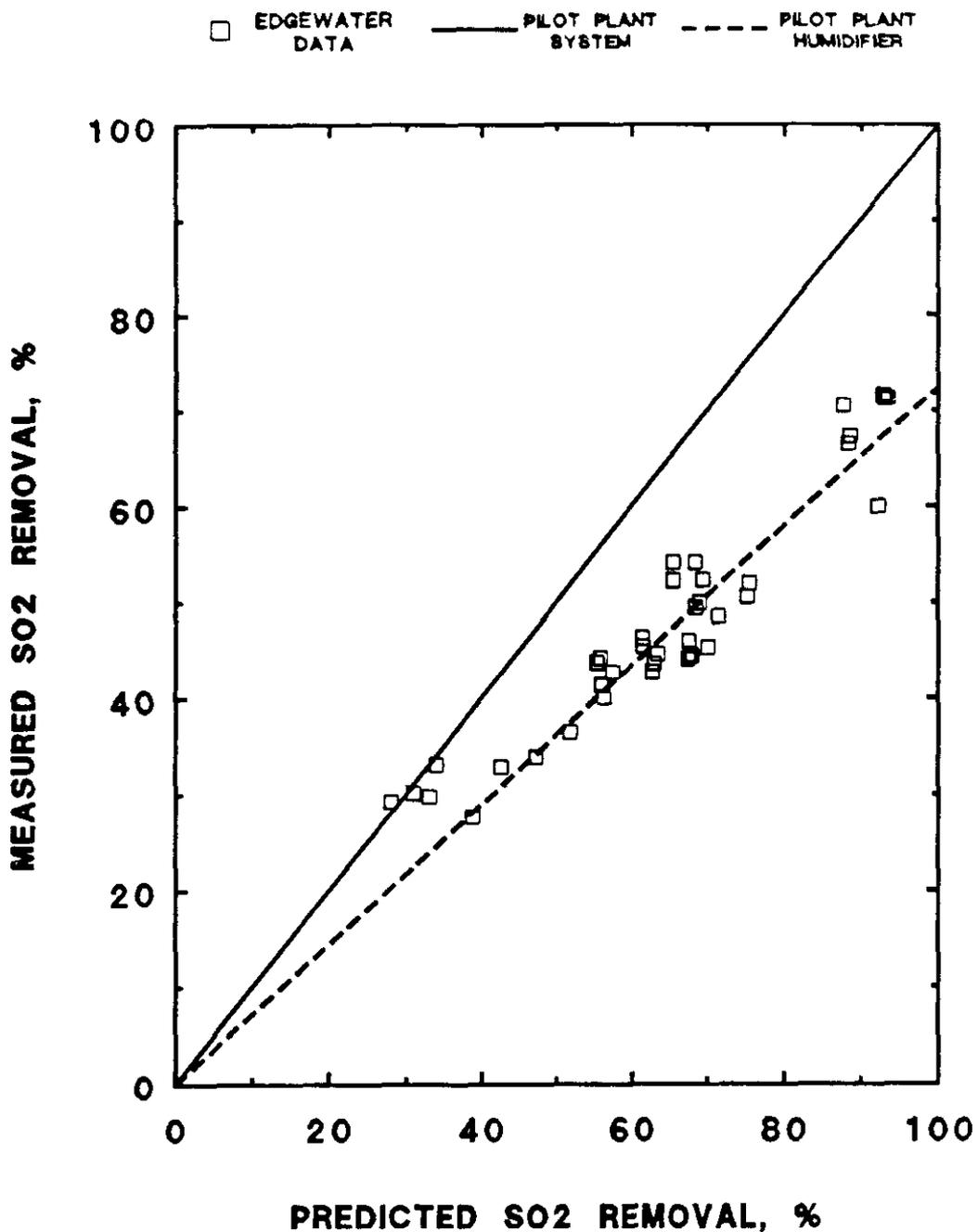


Figure 16. Comparison of pilot plant data with Edgewater data using hydrated lime A at 20 to 27°F approach to adiabatic saturation temperature, 0 to 0.24 Na/Ca mol ratio. See text for explanation of figure.

system (humidifier + baghouse) as a function of the Ca/S and Na/Ca molar ratios using hydrated lime A.<sup>1</sup> These equations were used to compare the pilot plant results with the Edgewater results. Figure 16 shows the SO<sub>2</sub> removals measured at Edgewater using hydrated lime A plotted as a function of the system SO<sub>2</sub> removals predicted using the pilot plant correlation equation. Edgewater data from once-through tests at 0.75 to 2.25 Ca/S, 0.0 to 0.2 Na/Ca, and 20 to 27°F approach without recycle were used in the figure. These were the variable ranges for the tests used to develop the regression equation. There are some differences in the secondary variables between the Edgewater tests and the regression equation. In the Edgewater tests, the inlet temperatures were 260 to 290°F and the inlet SO<sub>2</sub> contents were 700 to 1850 ppm; the regression equation is based on a 300°F inlet temperature and 1600 ppm SO<sub>2</sub> content. The solid parity line in the figure represents a perfect correlation with system SO<sub>2</sub> removal. Most of the data points (squares) in Figure 16 fall below this line, indicating that total system (humidifier + ESP) SO<sub>2</sub> removal was lower at Edgewater than in pilot plant (humidifier + baghouse) tests at equivalent Ca/S and Na/Ca ratios. The dotted line in the figure represents the correlation of pilot plant SO<sub>2</sub> removal in the humidifier only. The Edgewater data are clustered around this line, indicating that the system SO<sub>2</sub> removals at Edgewater match the pilot plant humidifier removals. This should not be interpreted to mean that little or no SO<sub>2</sub> removal occurred in the ESP at Edgewater. Recall that the results of bypass tests described earlier showed that roughly 5 to 15% (absolute) SO<sub>2</sub> capture might have taken place in the ESP at Edgewater during non-bypass operation. The best conclusion that can be made at this time is that the humidifier and ESP provided less SO<sub>2</sub> capture than the pilot plant humidifier and baghouse.

Similar behavior was observed for data from tests using hydrated lime G. In the pilot plant tests, least-squares regression equations were developed for hydrated lime G as a function of Ca/S, Na/Ca, approach to saturation, inlet gas temperature, and inlet gas SO<sub>2</sub> concentration. The regression for lime G used more variables than for lime A, because the pilot tests were performed using lime G at more widely varying process conditions. Figure 17 shows the SO<sub>2</sub> removals measured at Edgewater using hydrated lime G as a function of the system SO<sub>2</sub> removals predicted using the correlation equation developed in the pilot tests. Edgewater data from once-through tests at 0.75 to 2.25 Ca/S, 0 to 0.2 Na/Ca and

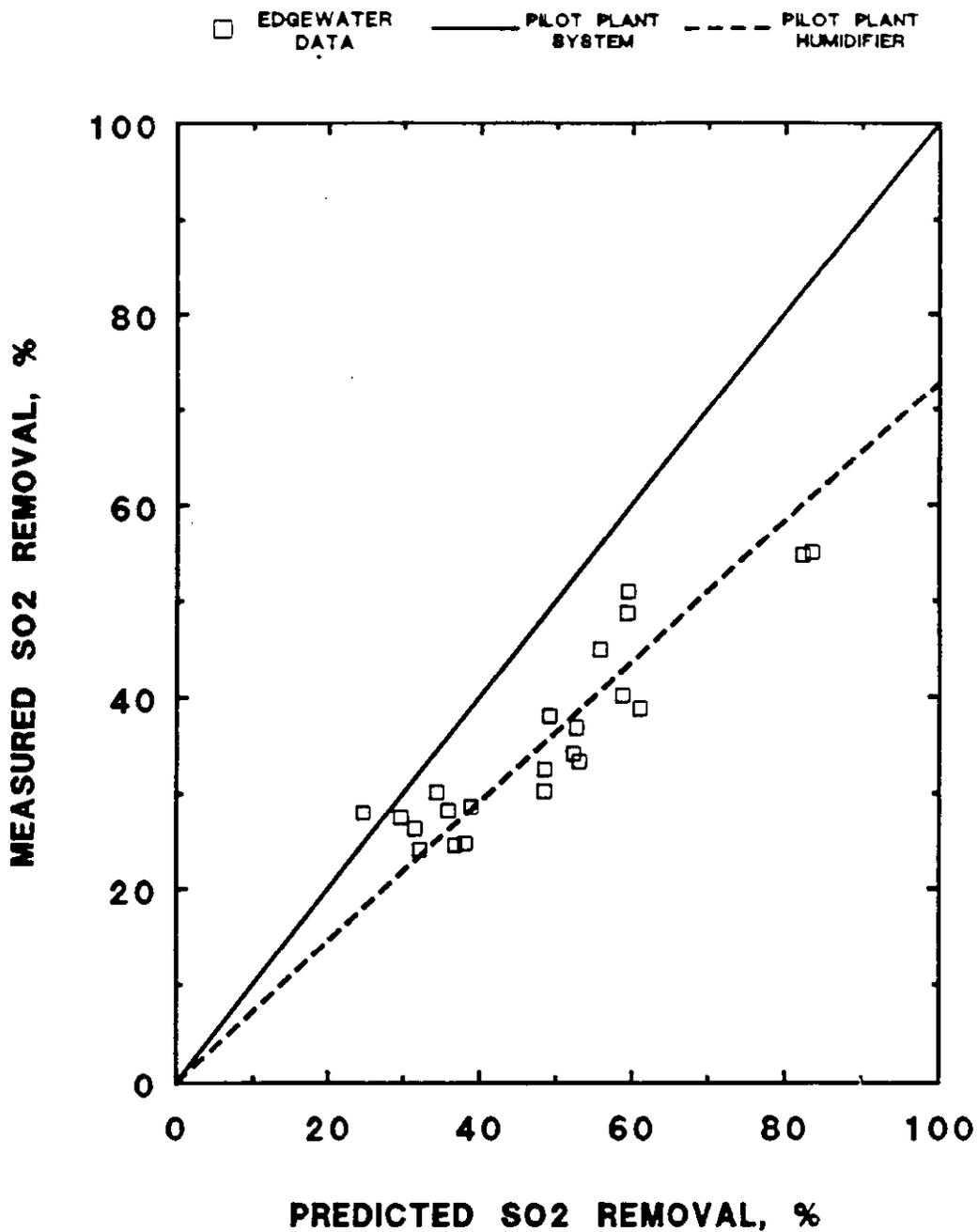


Figure 17. Comparison of pilot plant data with Edgewater data using hydrated lime G. See text for explanation of figure.

23-33°F approach were used in Figure 17. The humidifier inlet temperature and flue gas SO<sub>2</sub> concentration data were included in the calculation of predicted removals in Figure 17. Again, the solid parity line in the figure represents the correlation with system SO<sub>2</sub> removal. A few of the data points fall on or near this line at around 25 to 35% SO<sub>2</sub> removal, but most of the points are clustered around the dotted line representing SO<sub>2</sub> removal in the pilot plant humidifier. These results, which are similar to the results using hydrated lime A, indicate lower system SO<sub>2</sub> removals at Edgewater than those observed in the pilot plant at similar conditions.

There are many reasons why the system SO<sub>2</sub> removals were higher in pilot plant tests than in Edgewater tests at similar conditions. The most obvious difference between the two systems is the particulate collection device. Other differences between the two systems existed, including nozzle size and arrangement, humidifier design, and hydrated lime distribution in the flue gas.

The water spray nozzles were arranged in a 10 x 10 square array at Edgewater. In the pilot plant, a single spray nozzle was located in the duct center, which is a cylindrically symmetric system. The differences in the nozzle system designs may have resulted in significantly different coverage of the duct cross-sectional area by the water sprays and, thus, the interactions between lime particles and water droplets.

The lime injector positions, relative to the water spray atomizer positions, also were different. In the pilot plant, the hydrated lime was injected into the gas upstream of the water spray nozzle. At Edgewater, the hydrated lime was injected using five injectors located at the same plane as the water spray. The original injector design installation consisted of nine injectors in a 3 x 3 array to provide a relatively uniform particle distribution concentrated in the spray. However, due to transport air flow limitations and excessive floor deposits, only five of the injector ports could be used at once. The injector ports chosen were the three top row ports and two outside ports of the second row of the 3 x 3 array. This non-symmetric pattern probably did not give as uniform a particle distribution as in the pilot tests, since for the Edgewater tests most

of the solids were injected in the upper portion of the spray area. Again, different droplet-particle interactions may have developed.

Finally, a 20°F approach set point in the pilot plant may not represent the same humidity conditions in the two systems. At Edgewater it was more difficult to get an accurate representation of the wet and dry bulb temperatures in the large 14'7" x 14'7" duct. The wet bulb temperature normally was a single-point measurement taken every four to six hours in the duct center, and the outlet gas temperature was based on the average reading from five thermocouples. In the pilot plant, the thermocouple in the center of the 8.3-inch diameter duct provided a more accurate representation of the gas temperature, and the wet bulb temperature was measured every 30 minutes.

The effects of these differences, if any, on SO<sub>2</sub> removal have not been quantified.

#### FUTURE DIRECTIONS FOR PROCESS DESIGN AND PERFORMANCE IMPROVEMENTS

The Edgewater SO<sub>2</sub> removal data do not represent optimized process performance. The demonstration testing did not allow possible performance improvement through optimization of the process design, particularly the hydrated lime distribution in the flue gas at or just ahead of the humidification water atomizers in the humidifier duct. Sorbent recycle is another obvious option for process performance improvement that needs to be more fully evaluated. Further, other means of improving the process performance are possible. Process sorbent performance improvement is desirable to make the Coolside process more attractive for retrofit applications, since the demonstrated sorbent utilization is relatively low at 35% or less. Any improvement in the sorbent utilization will provide significant positive impact on the process economics by reducing the sorbent supply and resultant solid waste.

In the demonstration tests, hydrated lime was fed to highly localized areas, using five injectors, as discussed in the previous section. The initial design included nine injectors covering a wide duct cross-sectional area, but all of the injectors could not be used simultaneously because of design transport air flow

limitations (see discussion under "Test Program History" later in this section and also Section 4). Potentially, the SO<sub>2</sub> removal can be improved by improving the sorbent distribution over what was achieved during the Edgewater demonstration, since increased interactions of sorbent particles with water droplets are important for humidifier SO<sub>2</sub> capture. This possibility should be evaluated in the future by performing the process tests employing different sorbent dispersion patterns.

In the demonstration tests, sorbent recycle showed positive effects on the sorbent utilization. However, the tests were not sufficiently extensive to quantify conclusively the utilization improvement for long-term steady-state conditions.

Design of a humidifier to operate at closer approaches to adiabatic saturation will improve sorbent utilization. However, as approach temperature decreases, process operability becomes more of a concern because of the potential for forming wet sticky solids. Tighter temperature control will be required, and a reliable continuous wet bulb measurement device is a key need for closer approach operation.

For commercial Coolside process applications, additive costs will be a larger concern than for the demonstration program. Because of this, Na<sub>2</sub>CO<sub>3</sub> is preferred to NaOH for sorbent activation. Pilot plant tests indicate Na<sub>2</sub>CO<sub>3</sub> should activate the sorbent to the same extent as NaOH when used on an equimolar Na basis. Full-scale testing with Na<sub>2</sub>CO<sub>3</sub> is desired to confirm the pilot data.

In the longer term, optimization of the sorbent (hydrated lime) properties for SO<sub>2</sub> capture is expected to lead to an improved sorbent. Pilot plant tests have shown a positive correlation of hydrated lime surface area with sorbent utilization. Lime hydration methods that produce high surface area hydrates are being studied at Conso1 R&D<sup>7</sup> and elsewhere.<sup>8</sup> Additive incorporation during lime hydration also may provide more reactive sorbents.<sup>9,10</sup>

## DEMONSTRATION TEST METHODS

### Test Coals and Hydrated Limes

During the Edgewater tests, the boiler was fired with either a compliance (1.4 wt % S) or a non-compliance (3 wt % S) bituminous coal from Ohio. Typical coal analyses are given below. A table listing the complete analyses of all coal samples is included in Appendix A.

Coal	Btu/lb	Proximate Analysis			Ultimate Analysis					
		Mois- ture	Volatile Matter	Fixed Carbon	Ash	C	H	N	S	(by diff.)
Compliance	13204	4.18	34.75	54.74	10.51	74.48	4.92	1.39	1.42	7.29
Non-Compl.	12695	4.12	37.98	48.91	13.11	70.72	4.88	1.25	3.02	7.02

All analyses except moisture are dry wt % basis.

Two high-calcium hydrated limes were used. These were designated A and G which also were the designations used in pilot plant screening tests.<sup>5</sup> In the pilot screening tests, a total of ten hydrated calcitic and two pressure-hydrated dolomitic limes were evaluated to select the sorbents to be tested at Edgewater. The limes screened were all supplied by commercial producers. Hydrated lime A had the highest SO<sub>2</sub> removal activity in pilot plant tests and hydrated lime G was the lowest cost, based on the price delivered to the site. Both are high in calcium (>88% Ca(OH)<sub>2</sub>), but lime G has about three times the magnesium content as lime A. Lime A has a higher surface area (23 versus 17 m<sup>2</sup>/g) and calcium hydroxide content (93 versus 88 dry wt %) than lime G. Typical analyses are listed in Table 2. Tables listing the complete analyses of all lime samples are included in Appendix A. A third hydrated lime, hydrated lime H, was used without sodium addition during shakedown testing. This lime had been used in the LIMB/humidification tests which were performed at Edgewater prior to the Coolside test program. The data from these tests are not included in the discussion of desulfurization performance because of the limited number of runs made (ten).

## Data Measurements, Analyses and Work-Up Methods

Most of the data were collected using the Babcock & Wilcox System 140™ data collection system. These data were mainly temperatures, gas concentrations, or flow rates. A description of the major process measurements and calculation methods follows.

### Gas Concentrations--

The concentrations of SO<sub>2</sub> and O<sub>2</sub> were measured on a wet basis at the humidifier inlet and on a dry basis at the ESP exit using continuous gas analyzers. ThermoElectron SO<sub>2</sub> and Westinghouse O<sub>2</sub> analyzers were used at the humidifier inlet in a sampling system supplied by Enviroplan. A Du Pont SO<sub>2</sub> analyzer was used at the ESP exit for the first two months of the test program. For the last four months of operation, a Western SO<sub>2</sub> analyzer was used. A Thermo O<sub>2</sub> analyzer was used at the ESP exit. The gas analyzers were calibrated once a day using standard gas.

### Gas Temperatures--

Gas temperatures were measured at the humidifier inlet, humidifier exit, and ESP inlet using five unshielded thermocouples at each location arranged in a square-plus-center-point configuration (Figure 18). There were also five shielded thermocouples at the humidifier outlet. The humidifier outlet thermocouples were positioned immediately in front of the humidifier exit turning vanes. The outlet shielded thermocouples also were initially used as process thermocouples for controlling the humidifier outlet temperature. However, after about 2.5 months of testing (the week of October 20-25, 1989), an additional set of unshielded thermocouples was installed immediately after the exit turning vanes (called "turning vane" thermocouples) and process control was switched from the shielded outlet thermocouples to the turning vane thermocouples. This was done because of deposits that sometimes built up on the outlet thermocouples.

### Adiabatic Saturation (Wet Bulb) Temperature--

This temperature was taken at least once every eight hours using a thermocouple wrapped with a wet cotton wick. The thermocouple was at the end of a seven-foot probe. The probe was inserted into the center port at the humidifier exit just in front of the humidifier turning vanes. The wick was

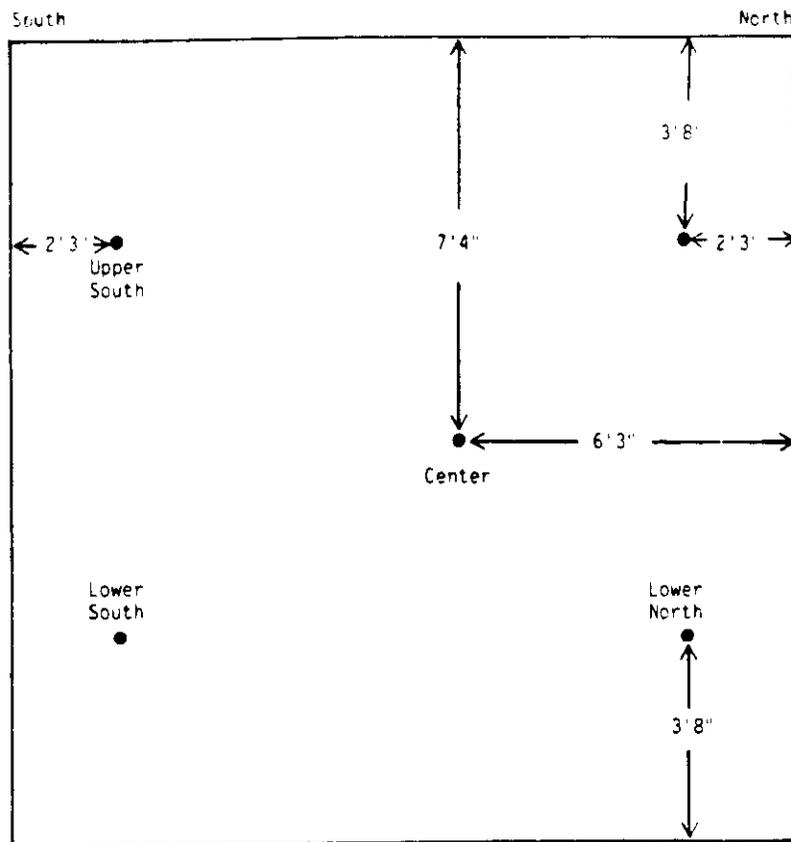


Figure 18. Arrangement of gas thermocouples at humidifier exit at Edgewater. Duct dimensions are 14'7" x 14'7".

immersed in water before inserting the probe into the duct. A hand-held digital readout unit was used to read the thermocouple temperature. The wet-bulb temperature was determined using the following procedure. Immediately after insertion, the temperature would rise as the hot gas warmed the water in the wick. As the water began to evaporate, the rate of temperature rise would level off due to the cooling effect of the evaporating water. When the heat input from the hot gas was balanced by the evaporative cooling by the water, the temperature reading became steady for one to three minutes. When water evaporation was nearly complete, the temperature would again rise to the gas (dry bulb) temperature. The reading during the one to three minute steady period was recorded as the wet bulb temperature, or the adiabatic saturation temperature. This procedure was complicated somewhat by solids (sorbent and fly ash) that were entrained in the gas depositing on the wick. To prevent a heavy build-up, the thermocouple probe was tapped and/or shaken every five to ten seconds while the probe was in the duct to knock off the solids. A wet bulb temperature also was taken at the humidifier inlet, ahead of the lime injectors and water spray, as a check. Generally the agreement was good, with the wet bulb temperature at the humidifier inlet reading within one to two degrees of the reading at the humidifier exit. If the difference was more than two degrees, the wet bulb measurements were repeated. The accuracy of the method was estimated to be  $\pm 2^\circ\text{F}$ , based on the typical accuracy of thermocouples ( $\pm 1$  to  $2^\circ\text{F}$ ) and the hand-held readout ( $\pm 1^\circ\text{F}$ ).

For about a third of the tests, the wet bulb temperature was not measured during the test run. For data analysis on these tests, the wet bulb temperature was estimated, based on a statistical correlation developed using the measured wet bulb temperatures and the temperatures and oxygen contents of the inlet flue gas during two different periods. For the October 2, 1989 to January 29, 1990 test period, the correlation is as follows:

$$\text{W.B.} = 140.94 - 13.66 \left( \frac{20.9}{20.9 - \text{O}_2\%} \right) + 0.0108 T_{\text{IW}} \quad (7)$$

For the February 1 to February 16, 1990, test period, the correlation is as follows:

$$\text{W.B.} = 144.29 - 17.00 \left( \frac{20.9}{20.9 - O_2\%} \right) + 0.0108 T_{IN} \quad (8)$$

where:

W.B. = estimated wet bulb temperature (°F)

O<sub>2</sub>% = vol % O<sub>2</sub> in the flue gas, measured at the humidifier inlet

T<sub>IN</sub> = flue gas<sup>2</sup> temperature measured at the humidifier inlet (°F)

The coefficient of determination (R<sup>2</sup>) is 0.60 for the first equation (166 data points) and 0.63 for the second equation (61 data points). The wet bulb temperature generally increases with increasing gas (dry bulb) temperature, and the oxygen concentration is a measurement of excess air dilution, which decreases the wet bulb temperature. Other factors that may affect the wet bulb temperature (such as the ambient air moisture content) were not measured during each test run. Two separate time periods were used for the correlation development because a steam tube leak was repaired January 30-31, 1990. The wet bulb temperatures were lower by 1 to 2°F at similar temperatures and O<sub>2</sub> concentrations after the steam tube repairs, because less water vapor was present in the flue gas. The correlation equations match the measured wet bulb temperatures to within ±2.5°F with a 90% statistical confidence level.

For the process run data listed in Appendix A, approaches based on the estimated wet bulb are shown in parentheses while approaches based on the measured wet bulb are shown without parentheses.

#### Flue Gas Flow--

The humidifier inlet flue gas flow rate was measured using two thermal dispersion mass flow probes. Each probe contained three measuring units (Figure 19). From January 9 through 23, 1990, these were miscalibrated by 10%. The flue gas flow data for tests during this period were adjusted accordingly. The data listed in Appendix A are the adjusted data.

#### Hydrated Lime Feed--

These flow rates were measured with gravimetric-type solids feeders, which used a weight loss over time method to determine rate.

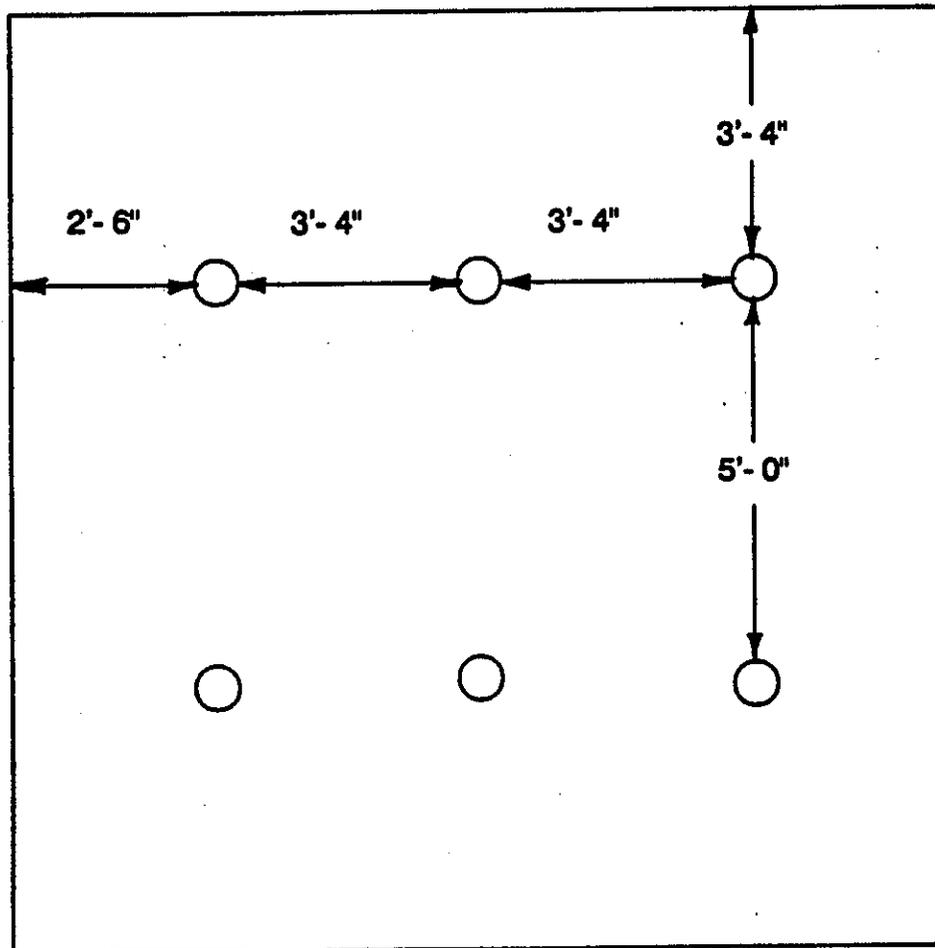


Figure 19. Arrangement of the flue gas flow meter detector locations in the duct at the humidifier inlet. Duct dimensions are 11'7"x11'7".

#### Humidification Water, Atomizing Air, NaOH Solution Flow--

The humidification water and NaOH flow rates were measured by in-line process magnetic-type mass flow meters. Atomizing air flow was measured by a vortex shedding meter.

#### Gross MW--

This was the total electric output of the plant's steam turbine generator, measured continuously by the power plant.

#### Coal Flow--

This was not measured directly, but instead was calculated by the Babcock & Wilcox System 140™ Diagnostic System, based on a heat balance using the total steam flow, the steam and flue gas temperatures and pressures, and the energy content of the coal.

#### Coal Sulfur Content--

The sulfur content of the coal samples was measured by Radian Corporation using a LECO total sulfur analyzer. The analysis was done on an as-received basis. The coal samples were taken every four to six hours using the following method. A cyclone sampler was permanently attached to each of the four coal transfer lines which connected the pulverizer mills to the burners. These samplers were usually in a constant purge state using high pressure nitrogen. A three- to five-minute sample from each sampler was taken simultaneously by shutting off the purge and allowing the positive pressure in the coal transfer lines to force the coal into the samplers. The four pulverizer samples were then combined into one composite sample, mixed thoroughly, and submitted to Radian's on-site lab for analysis. One daily composite sample was submitted to Commercial Testing & Engineering Company for proximate and ultimate analyses. However, since SO<sub>2</sub> removal measurements were based mostly on inlet and outlet gas analyzer data, the coal sulfur measurement was used to confirm the validity of the data.

#### Humidification Water NaOH Content--

Samples of the humidification water were taken from the water mixing and storage tank pump discharge line. These samples were taken every two to four hours to confirm the NaOH content. The samples were analyzed using a

conductivity meter and a standard curve of solution conductivity versus NaOH content. The conductivity meter was calibrated every day.

#### ESP Solids Samples--

These samples were taken for later analysis at Consol R&D. Samples were taken from one of the twelve ESP hoppers (Hopper 2A), using a two-inch pipe inserted near the bottom of the hopper. The other end of the pipe was connected to a hose leading to a 55-gallon sampling drum. Vacuum was drawn on the drum, hose, and sampling pipe using a high-pressure air eductor. Solids drawn through the pipe and hose collected in the drum until the drum was approximately 1/3 full. Of this, 100 to 200 g was submitted for analysis. Sampling time was usually less than one minute. Prior to sampling, the ESP operator was asked to empty the ESP hopper in the usual manner. When the hopper was empty, the hopper dump valve was shut and the hopper was allowed to fill for two to six hours before a sample was taken.

#### Data Analysis Procedures

The data analysis uses the raw process data generated by the System 140™ computer (which represents 10-minute averages of corresponding process data collected every minute during a given test period) and the daily average sulfur content of the coal (the average of the day's coal sulfur values as determined by Radian Corporation using a Leco analyzer). The first data reduction step is to calculate the values of important process variables (Table 9) from the raw process data. The calculation methods are detailed later. Based on trends in the calculated process variable values, the test period is chosen. This is usually done by examining graphs of the process variables versus time and choosing a time period over which each of the process variables appears to be constant. The minimum test duration is one hour. Next, each process variable value is averaged for the entire test period. The statistical standard deviation is also calculated for each process variable to check for stable conditions. Stable conditions are defined as having a standard deviation of less than 10% of the average value for each process variable. If the criterion for stable conditions is not met, another time period is selected when the standard deviations are all below 10%. An example of the variability of the tests is

TABLE 9  
IMPORTANT PROCESS VARIABLES

Average Humidifier Inlet Temperature, °F	
Average Humidifier Outlet Temperature (Shielded), °F	
Average ESP Inlet Temperature, °F	
Atomizing Air/Water Ratio, lb/lb	(Eq. 9)
Water/Flue Gas Ratio, gal/1000 lb	(Eq. 10)
SO <sub>2</sub> ppm Corrected to Dry, 0% Excess Air (Humidifier Inlet)	(Eq. 11)
SO <sub>2</sub> ppm Corrected to Dry, 0% Excess Air (Humidifier Outlet)	(Eq. 12)
System Ca/S, mol/mol Based on Coal Feed and % S	(Eq. 13)
Humidifier Ca/S, mol/mol Based on Humidifier Inlet Gas Analyzers	(Eq. 14)
NaOH/Ca(OH) <sub>2</sub> by Weight	(Eq. 15)
% SO <sub>2</sub> Removal Based on Coal Feed and % S (Early Tests)	(Eq. 16)
% SO <sub>2</sub> Removal Based on Humidifier Inlet Analyzer Data	(Eq. 17)
Baseline SO <sub>2</sub> Content (ppm) from % S in Coal	(Eq. 18)

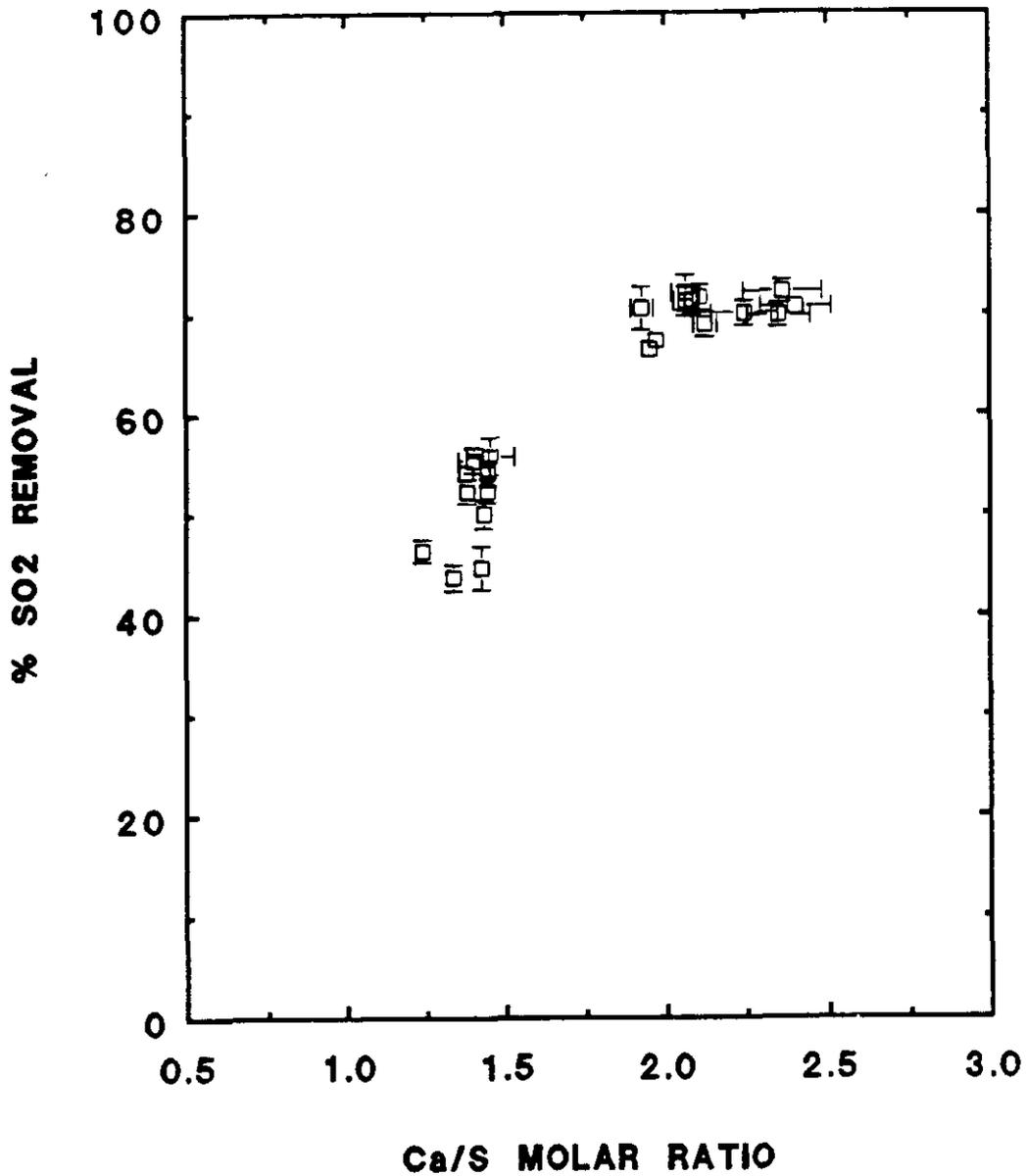


Figure 20. Variability of SO<sub>2</sub> removal and Ca/S ratio for tests using hydrated lime A at 0.17 to 0.24 Na/Ca and 19 to 22°F approach. Error bars represent one standard deviation. Points without error bars have error smaller than plot symbol.

shown by the error bars in Figure 20, which represent standard deviations in the SO<sub>2</sub> removal and the Ca/S ratio during each test run period using hydrated lime A at 0.2 Na/Ca and 20°F approach. The data points in Figure 20 are the same data as the circles in Figure 1.

The following calculations are performed using the archived raw process data collected by the System 140™ computer and daily average % S content to obtain the values of the process variables used for data analysis as discussed above.

### Process Variable Calculation Methods

**Average Gas Temperatures:** The readings of five thermocouples in an array (see Figure 18 for the location of the TCs) are averaged. This calculation is performed for three locations: humidifier inlet, humidifier outlet, and ESP inlet.

Atomizing Air/Water Ratio, lb/lb:

$$\frac{\text{Atomizing Air Flow, lb/hr}}{\text{Humid Water Flow gpm} \times 60 \text{ min/hr} \times 8.34 \text{ lb/gal}} \quad (9)$$

Water/Flue Gas Ratio, gal/klb:

$$\frac{\text{Humid Water Flow, gpm} \times 60 \text{ min/hr}}{\text{Measured Humidifier Inlet Gas Flow, lb/hr}} \quad (10)$$

SO<sub>2</sub> ppm Corrected to Dry, 0% Excess Air (Humid Inlet):

$$1 - \frac{\text{Humid Inlet SO}_2 \text{ ppm (wet)}}{\text{Humid Inlet O}_2 \% \text{ (wet)}} = 0.07 \quad (11)$$

20.9

(Note: The moisture content of the flue gas was assumed to be 7%.)

SO<sub>2</sub> ppm Corrected to Dry, 0% Excess Air (ESP Outlet):

$$1 - \frac{\text{ESP Outlet SO}_2 \text{ ppm (dry)}}{\frac{\text{ESP Outlet O}_2 \text{ \% (dry)}}{20.9}} \quad (12)$$

System Ca/S mol Ratio, Based on Coal Feed and % S:

$$\frac{\text{Hydrated Lime Feed Rate, klb/hr}}{(\text{Daily Avg. \% S in Coal})(\text{Coal Feed Rate lb/hr})} \times \quad (13)$$

$$\frac{32 \text{ lb S/mol}}{74.1 \text{ lb hydrated Lime/mol}} \times \frac{\% \text{ Lime Purity}}{1000 \text{ lb/klb}}$$

Humidifier Ca/S mol Ratio, Based on Humid Inlet Gas Analysis:

$$\frac{\text{Hydrated Lime Feed Rate, lb/hr}}{(\text{Humid Inlet SO}_2 \text{ ppm wet})(\text{Humid Gas Inlet Flow, lb/hr})} \times \quad (14)$$

$$\frac{29.5 \text{ lb Flue Gas/mol}}{74.1 \text{ lb Lime/mol}} \times \frac{\% \text{ Lime Purity}}{1/(10^6 \text{ ppm})} \times \frac{1}{100\%}$$

(Note: Lime purity was 93% for Lime A and 88% for Lime G, based on average TGA analysis.)

NaOH/Ca(OH)<sub>2</sub> Ratio by Weight:

$$\frac{(\text{NaOH Solution Flow Rate, lb/hr})(\% \text{ NaOH Purity})}{(\text{Hydrated Lime Feed Rate, lb/hr})(\% \text{ Lime Purity})} \quad (15)$$

(Note: Lab analyses showed an average NaOH solution purity of 47% by weight.)

% SO<sub>2</sub> Removal Based on Coal Feed and % S:

$$\left( 1 - \frac{\text{ESP Outlet ppm SO}_2, \text{ Corrected to Dry, 0\% Excess Air}}{\text{Baseline* SO}_2 \text{ ppm from \% S in Coal}} \right) * 100\% \quad (16)$$

% SO<sub>2</sub> Removal Based on Humid Inlet Gas Analysis:

$$\left( 1 - \frac{\text{ESP Outlet ppm SO}_2, \text{ Corrected to Dry, 0\% Excess Air}}{\text{Humid Inlet ppm SO}_2, \text{ Corrected to Dry, 0\% Excess Air}} \right) * 100\% \quad (17)$$

\*Baseline SO<sub>2</sub> (ppm) from % S in Coal: This is the SO<sub>2</sub> ppm (dry, 0% excess air) calculated from the daily average % S in the coal. The EPA F-number method is used to calculate the flue gas dry volume. The calculation is:

$$\frac{\text{Daily Average \% S in Coal}/100)(2 \text{ lb SO}_2/\text{lb S})}{(\text{x dscf/lb Coal})(0.1689 \text{ lb/scf SO}_2)} \times 10^6 \text{ ppm} \quad (18)$$

(Note: The value x dscf/lb coal is based on the coal analysis.)

### Test Program History

Coolside tests were performed from July 31, 1989, through February 16, 1990. Eleven sets of tests were performed using three different hydrated limes with and without sodium additive and recycle. Several unplanned process changes were made during the course of the test program, including sorbent injector outlet repositioning, the installation and use of another set of gas temperature thermocouples for process control, flue gas bypass operation at high load conditions, and a 5°F increase in the approach to adiabatic saturation temperature set point from 20°F to 25°F. Table 10 lists the tests performed in chronological order with the test dates and hydrated limes used.

After the initial start-up and shakedown test period, the first ten Coolside tests were performed to select a sorbent injector configuration for later tests (Test Set I). The boiler was fired with a compliance coal. Hydrated lime H, left over from the LIMB/Humidification tests which were performed prior to Coolside start-up, was the sorbent used. NaOH additive was not used. Two injector configurations were tested: the first made a wide lime dispersion across the duct cross-section and the second made a lime dispersion more concentrated over the duct area covered by the humidification water nozzles. Both

TABLE 10  
SCHEDULE OF COOLSIDE TESTS AT EDGEWATER

<u>Date</u>	<u>Activity</u>
<u>Test Set I - Hydrated Lime H</u>	
7/31 - 8/17/89	Sorbent Injector Selection Tests: No additive, compliance coal, 20°F approach
8/18 - 8/27/89	Boiler down for maintenance repairs
<u>Test Set II - Hydrated Lime A</u>	
8/28 - 9/28/89	Once-Through Tests: No additive, compliance coal, 20°F approach
9/29 - 10/1/89	Additive delivery system start-up and shake-down
<u>Test Set III - Hydrated Lime A</u>	
10/2 - 10/19/89	Once-Through Tests: NaOH injection, high sulfur coal, 20°F approach
10/20 - 10/25/89	Turning vane thermocouples installed; flue gas bypass capability added to system controller
<u>Test Set IV - Hydrated Lime A</u>	
10/26 - 11/2/89	Once-Through Tests: NaOH injection, compliance coal, 20°F approach
11/3 - 11/10/89	Boiler down due to problem with plant water intake
<u>Test Set V - Hydrated Lime A</u>	
11/11 - 11/15/89	Once-Through Tests: NaOH injection, high sulfur coal, 25°F approach
11/16 - 11/27/89	Thanksgiving - no tests

TABLE 10 (Continued)  
 SCHEDULE OF COOLSIDE TESTS AT EDGEWATER

<u>Date</u>	<u>Activity</u>
<u>Test Set VI - Hydrated Lime A</u>	
11/28 - 12/14/89	Recycle Tests: Compliance coal, NaOH variable due to NaOH pump problems, 25°F approach
12/20/89 - 1/4/90	Christmas holiday - no tests
<u>Test Set VII - Hydrated Lime A</u>	
1/5 - 1/9/90	Once-Through Tests: No additive, compliance coal, 25°F approach
1/10 - 1/15/90	Gas analyzer breakdown - no data available. Humidifier was operated to run out supply of lime A in preparation for tests using lime G.
<u>Test Set VIII - Hydrated Lime G</u>	
1/16 - 1/19/90	Once-Through Tests: No additive, compliance coal, 25°F approach
<u>Test Set IX - Hydrated Lime G</u>	
1/20 - 1/24/90	Once-Through Tests: NaOH injection, compliance coal, 25°F approach
<u>Test Set X - Hydrated Lime G</u>	
1/25 - 2/3/90	Once-Through Tests: NaOH injection, high sulfur coal, 25° approach
<u>Test Set XI - Hydrated Lime G</u>	
2/5 - 2/16/90	Recycle Tests: NaOH injection, compliance coal, 25°F approach

injector sets had outlet impingement disks to help distribute the sorbent into the flue gases (see "Hydrated Lime Feed System" in Section 4). Some tests were made using the second injector set with distribution disks removed. The SO<sub>2</sub> removals were 17 to 18% at Ca/S of 0.9 to 1.0 and 25 to 48% at Ca/S of 1.7 to 2.0. The wide variation in SO<sub>2</sub> removal at 1.7 to 2.0 Ca/S was due to a wide range in the approach to adiabatic saturation, which was 17 to 23°F. No significant difference in SO<sub>2</sub> removals was observed between the two injector configurations under similar process conditions. However, because of the high degree of variability in the data, this does not represent a definitive test of the effect of sorbent distribution.

These results were not included in the data discussion section of this report, because only ten tests were performed using hydrated lime H. This was not enough to draw reliable conclusions regarding the effect of Ca/S or approach to saturation on SO<sub>2</sub> removal using this hydrated lime. During this test period, heavy lime deposition occurred on the humidification water atomizer lances. Otherwise, the humidifier operated well with only minor solid accumulation on the floor. Because the boiler was operated with a newly installed electronic control system which was being commissioned, boiler operation was erratic with occasional shutdowns taking place.

The next set of tests (Test Set II) consisted of once-through tests using hydrated lime A without additive. Compliance coal was fired in the boiler. The second solids injector set without the distribution caps was used in this set of tests because it was already in place. The injector location was originally 8 feet upstream from the water spray atomizers. During this period, heavy solids deposition on the water spray lances was resolved by extending the injector ports (on September 5, 1989) to the same vertical plane as the water spray atomizers. Because of transport air limitations, only five of the nine injector ports were used. Figure 21 shows the ports that were closed off. This reduced the deposition on the water spray lances. The results of Test Set II were not considered to be as reliable as the results of the subsequent tests because of recurring problems with the humidifier inlet gas analyzer. As an alternative, the ESP exit analyzer readings taken before and after the test were used to establish the

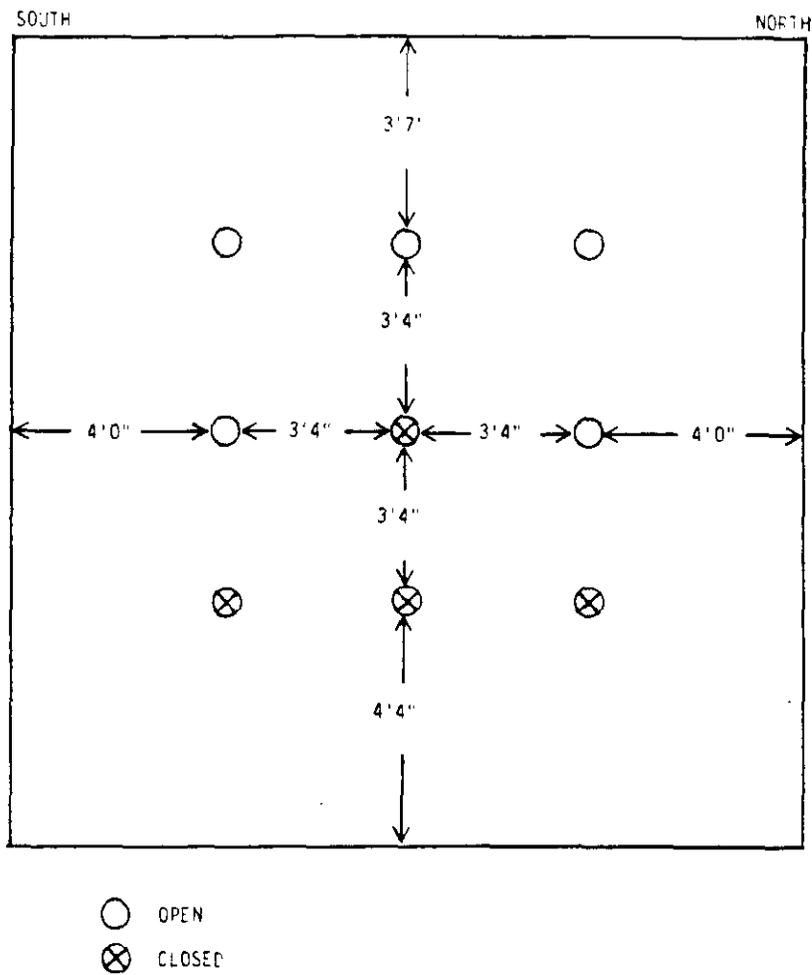


Figure 21. Arrangement showing lime injector port locations in the duct. Duct dimensions are 14'7" x 14'7".

baseline flue gas SO<sub>2</sub> levels. This did not account for baseline shifts that might have occurred due to coal sulfur content changes, for example.

Test Set III was a set of once-through tests performed using hydrated lime A with sodium hydroxide addition to the humidifier water. High-sulfur coal was fired in the boiler and the Coolside process kept the power plant in compliance with SO<sub>2</sub> emission standards. Demonstration testing was operated around the clock, and continued throughout the remainder of the test program. Previously, testing was not performed overnight. During the round-the-clock testing, heavy solids deposition occurred on the humidifier outlet turning vanes when the atomizing air-to-water ratio fell below the design level of 0.45 lb/lb. In subsequent tests, a portion of the hot flue gas was bypassed around the humidifier at high boiler load (>75 MW<sub>e</sub>) to reduce the water demand and keep the air-to-water ratio above 0.45 lb/lb to avoid the formation of large water droplets. Also, the ESP operation was modified to improve operation by using intermittent energization on the first two fields. An inspection of the ESP by Ohio Edison during this test period showed that the ESP was in good condition, except for some increase in wire deposits in the front fields (see "ESP Operation" in Section 4).

Test Set IV consisted of once-through tests performed using hydrated lime A with sodium hydroxide added to the humidification water. A new set of process control thermocouples was installed downstream of the turning vanes at the humidifier outlet, and flue gas bypass operation at high boiler load conditions was begun. Compliance coal was fired in the boiler to avoid out-of-compliance SO<sub>2</sub> emissions caused by the flue gas bypass. Since high boiler load operation usually occurred in the day time, most test data were obtained during the evening and night shifts when there was no bypass operation.

Test Set V was a continuation of Test Set III described earlier; however, the approach to adiabatic saturation set point was raised to 25°F beginning with these tests because of humidifier performance deterioration, which resulted in outlet turning vane deposits. The performance deterioration was shown by the unshielded outlet thermocouples, which were reading near-wet bulb temperatures at 20°F approach operation, indicating the presence of large water droplets

impinging on the thermocouples. An inspection of the nozzles showed that the atomizer holes had become larger due to erosion. Prior to this, the set point was 20°F approach.

Test Set VI focused on recycle tests, in which a portion of the solids collected by the ESP were recycled into the humidifier using injectors similar to the hydrated lime injectors. As with the fresh lime injection, the recycle solids injectors were located in the same vertical plane as the water spray atomizer nozzles. Fresh hydrated lime A and sodium additive also were used. Due to severe cold weather, the NaOH pump malfunctioned often and, thus, the use of NaOH additive was on a day-to-day basis during these tests. These tests continued until the Christmas holiday.

Test Set VII consisted of once-through tests without additive using compliance coal, the same as Set II, except the approach to saturation set point was 25°F instead of 20°F. These tests were performed because the NaOH pump was being repaired due to cold-weather damage. Compliance coal was used because SO<sub>2</sub> emissions compliance was not assured without NaOH injection. These were the last tests using hydrated lime A.

Hydrated lime G was used beginning with Test Set VIII. These were once-through tests without additive using compliance coal. All tests using hydrated lime G were performed at the 25°F set point approach to adiabatic saturation.

Test Set IX were once-through tests using hydrated lime G with sodium additive injection using compliance coal.

Test Set X were once-through tests using hydrated lime G with sodium additive injection using high-sulfur coal.

Test Set XI were recycle tests using recycle sorbent along with hydrated lime G with sodium additive. These tests are made using compliance coal. The final test during this set used recycle sorbent without fresh hydrated lime or sodium feeds. The recycle sorbent came from tests using hydrated lime G.

During Test Sets IX through XI, the NaOH pump broke down on occasion and additive feed was not continuous throughout. Thus, a few tests were performed during each set without additive.

## SECTION 4

### PROCESS RELIABILITY/OPERABILITY

#### HYDRATED LIME FEED SYSTEM

##### General Description

The major components of the sorbent feed system are shown in Figure 22. Hydrated lime was delivered by truck and pneumatically conveyed to the 300 ton hydrated lime storage silo. The hydrated lime was pneumatically transferred to a 20 ton capacity hydrated lime day bin, located inside of the boiler house, which supplied hydrated lime to a gravimetric feeder. Weigh cells located on the day bin I-beam supports were used to continuously measure the solids weight within the day bin. The day bin automatically was refilled when a low weight indication is received. The day silo filled until a high weight signal automatically halts hydrated lime transfer from the storage silo to the day bin. A baghouse located on top of the day bin filtered the conveying air which is vented.

The Coolside process testing required accurate measurement of the hydrated lime feed rate. A gravimetric rather than a volumetric feeder was used to supply the lime with the required accuracy. The gravimetric feeder consisted of a small weigh hopper and an integral variable speed feeder screw. The hopper weight loss over time determined the feed rate. The feeder control set point was input from the plant computer control system and was based on: boiler load, flue gas flow to the humidifier, coal sulfur content, and desired Ca/S ratio.

When the gravimetric feeder hopper weight dropped to low level, the hopper was automatically refilled from the day bin. The feeder control system automatically activates the day bin live bottom vibrator, started the rotary seal between the day bin and the gravimetric feeder and opened a slide gate valve below the day bin. Dust-containing air displaced from the hopper during filling was vented through a hose to the duct leading to the boiler air heater inlet. Venting of the feeder hopper was facilitated because the air heater duct is under vacuum. When the gravimetric feeder hopper was full, the slide gate valve closes

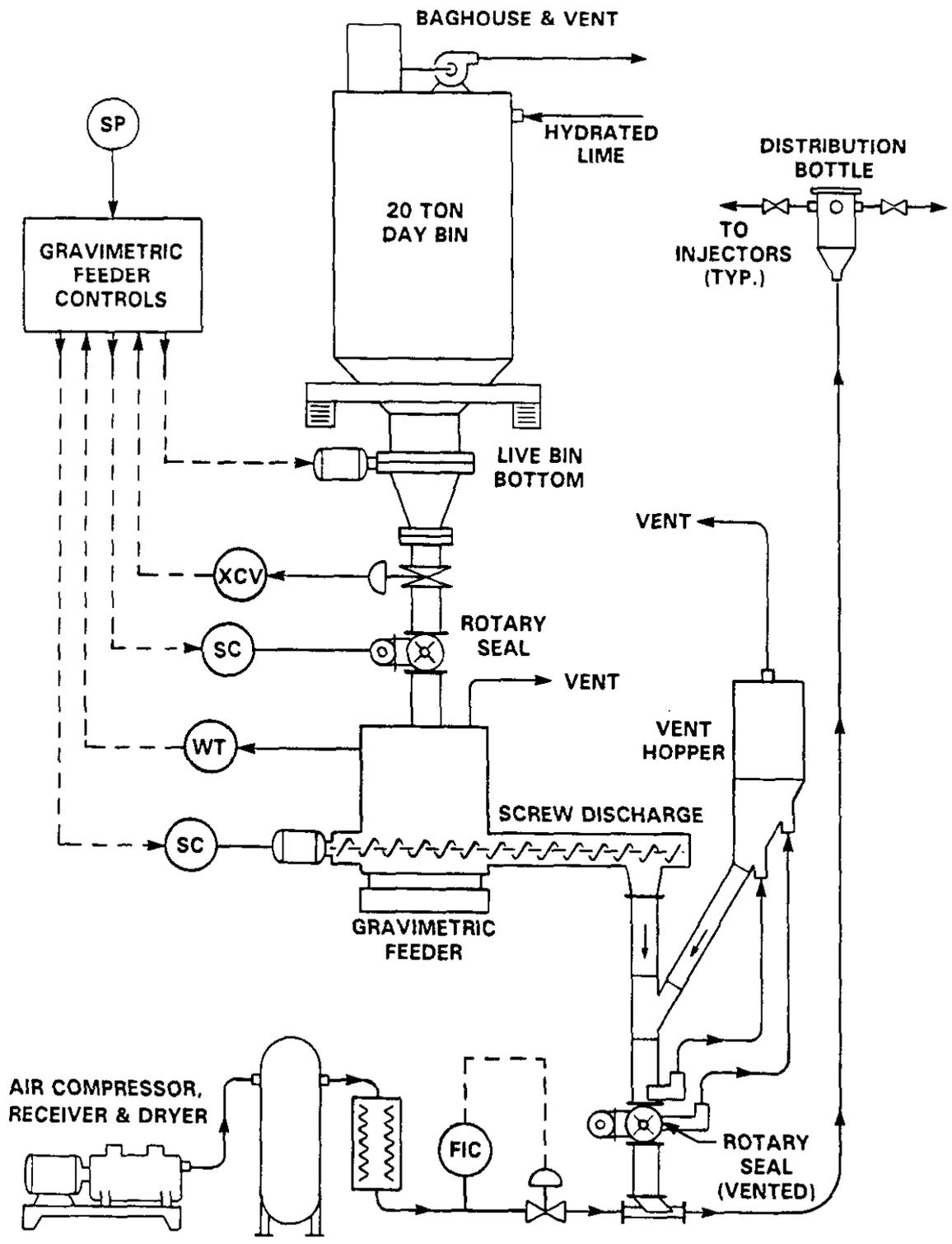


Figure 22. Edgewater Station hydrated lime feed system.

and the rotary seal and bin vibrator shut down. During the short 1-2 minute refill period, the feed screw speed was held constant.

The solids discharged from the gravimetric feeder screw drop through a rotary seal and were pneumatically conveyed to a solids distribution bottle. The distribution bottle was located on top of the humidifier. Up to nine solids-injector pipes were supplied from the distribution bottle. A dedicated compressor/dryer system supplied 900-1100 lb/hr of conveying air.

### Feeder Operation

Except for the distribution bottle and injector piping, the sorbent feed equipment used during the Coolside process tests was the same as that used in the preceding EPA LIMB test program. During the LIMB program, a number of equipment changes were made to improve operation. For example, a rotary seal was installed above the gravimetric feeder to prevent a too-rapid filling of the feeder weigh hopper. Rapid refilling of the hopper had on occasion caused fluidization of the solids remaining in the feeder hopper. This resulted in uncontrolled solids flushing through the gravimetric feeder screw into the sorbent supply lines.

Another modification was the installation of a rotary seal feeder pocket vent system. The system was needed to vent the conveying air which pressurized the empty rotary seal feeder pockets and which leaked past feeder vane seals. Before the vent system was installed, the sorbent feed rate was erratic due to the fluidization of the feeder hopper by conveying air which was not vented. To correct the problem, two vent reliefs were installed; one on the side of the rotary seal and one at the rotary seal inlet (see Figure 23). As shown in Figure 22, the feeder vents are connected to a vent hopper which allows separation of solids from the vented air. The solids return by gravity to the rotary seal inlet while the air vents through a rubber hose to the duct leading to the flue gas side of the air heater inlet which operates under vacuum.

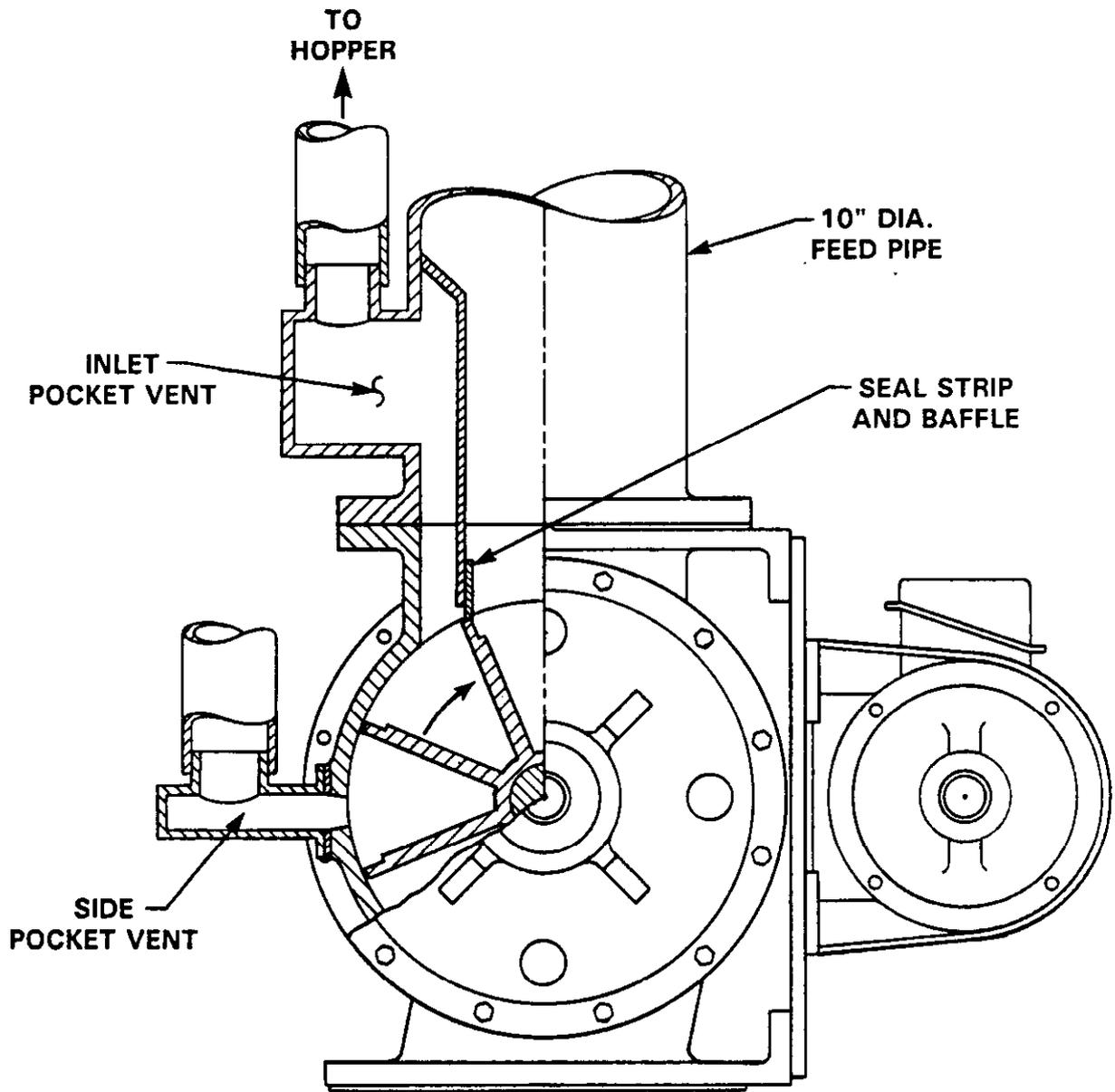


Figure 23. Modified rotary valve.

The vent system allowed adequate sorbent feed for the Coolside testing. Occasional feed problems occurred when dust plugged horizontal sections of the vent hopper pressure relief hose. This problem was minimized but not completely eliminated by re-routing the hose from the economizer outlet duct to the air heater flue gas outlet duct. The routing increased the vacuum on the hose and minimized the length of horizontal hose run. An improved rotary seal system using two seals in series or the installation of a powder pump solids feeder is recommended for commercial applications.

### Solids Injection Lance Design And Operation

Figure 24 shows the equipment for distribution and dispersion of hydrated lime and recycle solids into the flue gas at the humidifier inlet. Separate distribution bottles were used for the hydrated lime sorbent and for the recycle solids. Each feed stream can be split into nine separate streams. Rubber hoses connected the distribution bottles to the injector piping.

The injector pipes were originally fitted with short discharge nozzles as shown in Figure 25. The outlet distributor disk caused the pneumatically-conveyed solids to fan out from the nozzle perpendicular to the direction of flue gas flow. This distributed the solids in the flue gases ahead of the water spray lances. Two different injector piping arrangements were tested: one with a wide solids distribution pattern which was to spread the solids out uniformly across the entire duct and the other; and a closer packed arrangement which confined the solids distribution to the projected area of the water spray array. The operation of these injector arrangements is discussed further in Section 3 "Desulfurization Performance--Demonstration Test Methods--Test Program History."

A major operating problem became immediately apparent when hydrated lime was first fed to the system. The hydrated lime laid down rapidly on the water spray air foils and blocked off flow area between the water spray lances as shown in Figure 26. The material on the water lances was not a deposit but rather was settled dust. The flue gas passing through the water lances at 20 to 30 fps did not have sufficient momentum to carry away the dust buildups.

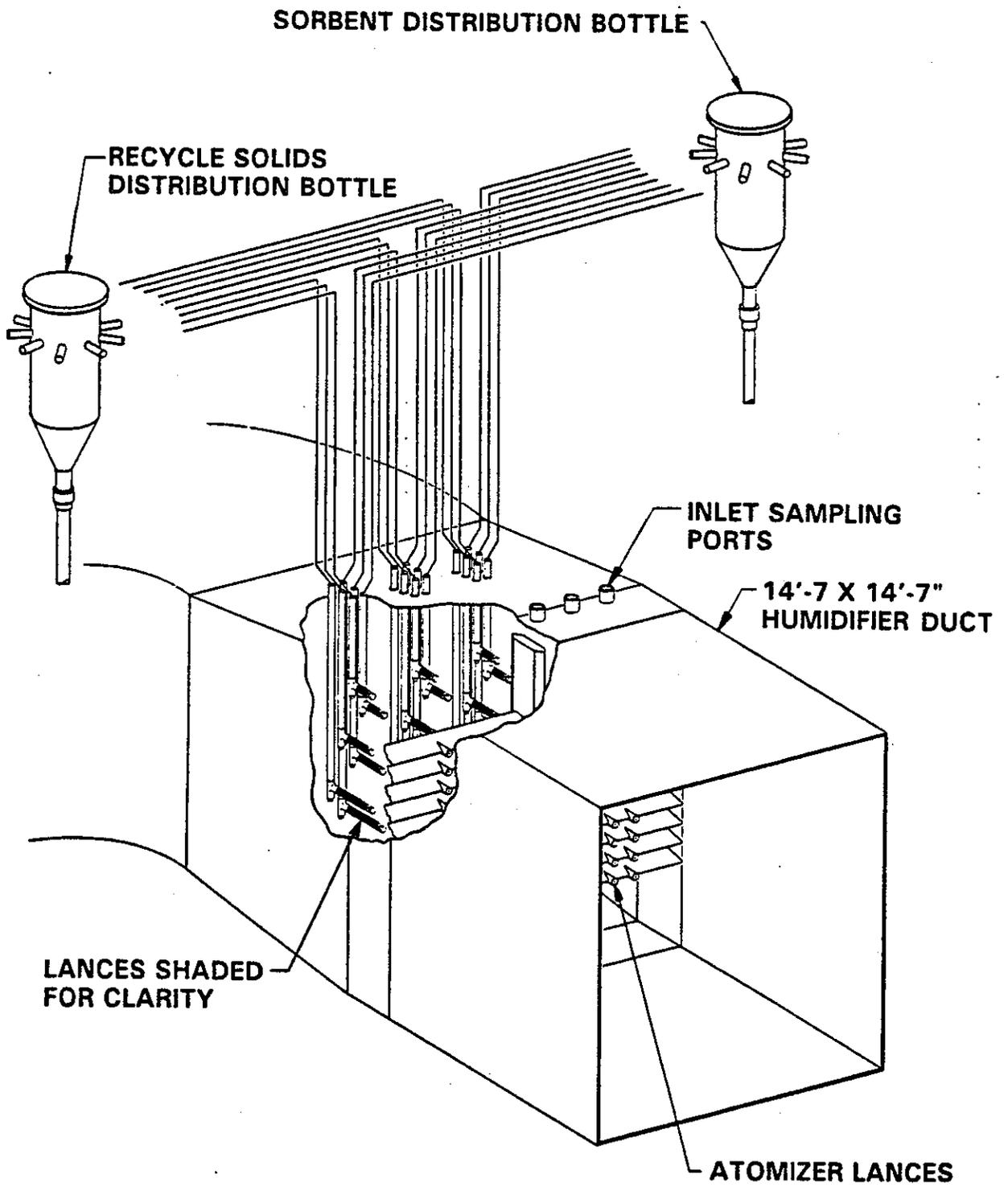


Figure 24. Original arrangement of short solids injection pipes.

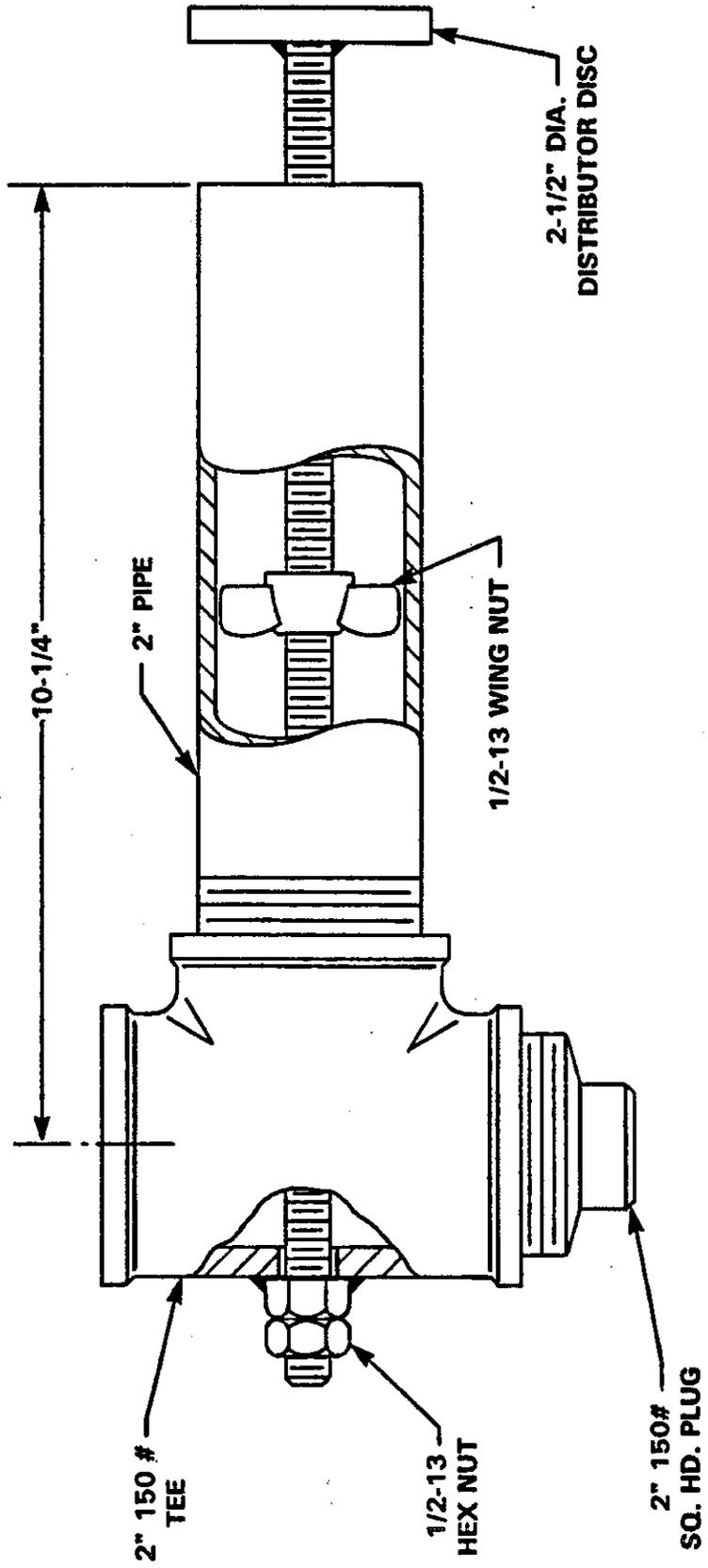


Figure 25. Original design of short solids injector pipes with outlet solids distributor disks.



Figure 26. Lime accumulation on water spray lances (backside view of water lances).

The physical character of hydrated lime appears to be responsible for the dust lay down. Hydrated lime solids do not fluidize and disperse as readily as solids like fly ash, but tend to clump and form particle agglomerates. Even at conveying air velocities of 30-35 fps, the hydrated lime solids discharging from the injector pipes were not uniformly mixed with the conveying air. The reason for this is not known. It may be due to lime particle surface moisture or electrostatic charge effects.

Initial attempts were made to remove the dust accumulations from the humidifier air foils by hand air lancing through humidifier inspection ports. This was ineffective because the dust accumulations would reform in 20 to 60 minutes. Another attempt was made to improve the dust removal by installing stationary air lances on each humidifier spray lance near the side walls and near the center support structure. The air lancing was automated. This proved ineffective since the air lances only cleared perhaps 10% of the water spray lance surface area which was just in front of the air blower tube outlets.

The problem was solved by extending the hydrated lime injection pipes through the water spray lances as shown in Figure 27. The hydrated lime discharged from the pipes at the plane of the water spray atomizers rather than upstream of the atomizers. To prevent gas flow interference with the water atomizer spray plumes, the outlet solids distribution disks (see Figure 25) were not installed on the injector pipes. The high turbulence created by the water sprays apparently distributed and mixed the lime with the flue gas to a degree since  $SO_2$  removals were similar to pilot-scale test results. Figure 28 shows that there were no dust accumulations on the water spray lances even after eleven days of continuous operation in which both hydrated lime and recycle solids were fed to the humidifier.

As indicated in Figures 24 and 27, there was a 3x3 array of hydrated lime and recycle solids injectors. The hydrated lime and recycle solids injector pipes were side-by-side but were slightly staggered vertically. For hydrated lime injection, normally only the top row and the side injectors of the second row were operated (five injectors total). The center injectors of the second row and the bottom row injectors were taken out of service. The center injectors were taken out of service because of solids buildups on four water spray

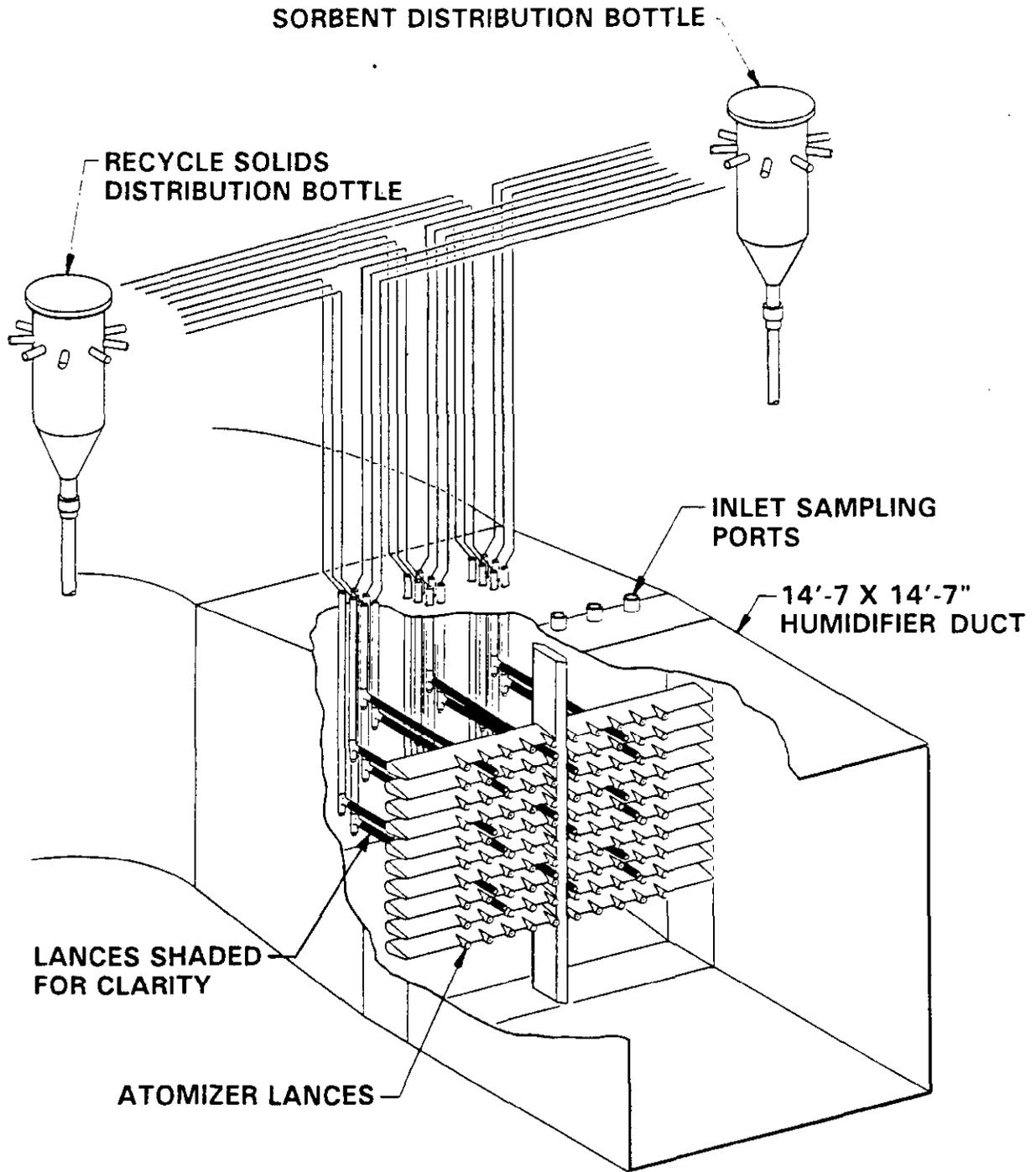


Figure 27. Final arrangement of solids injection lances.

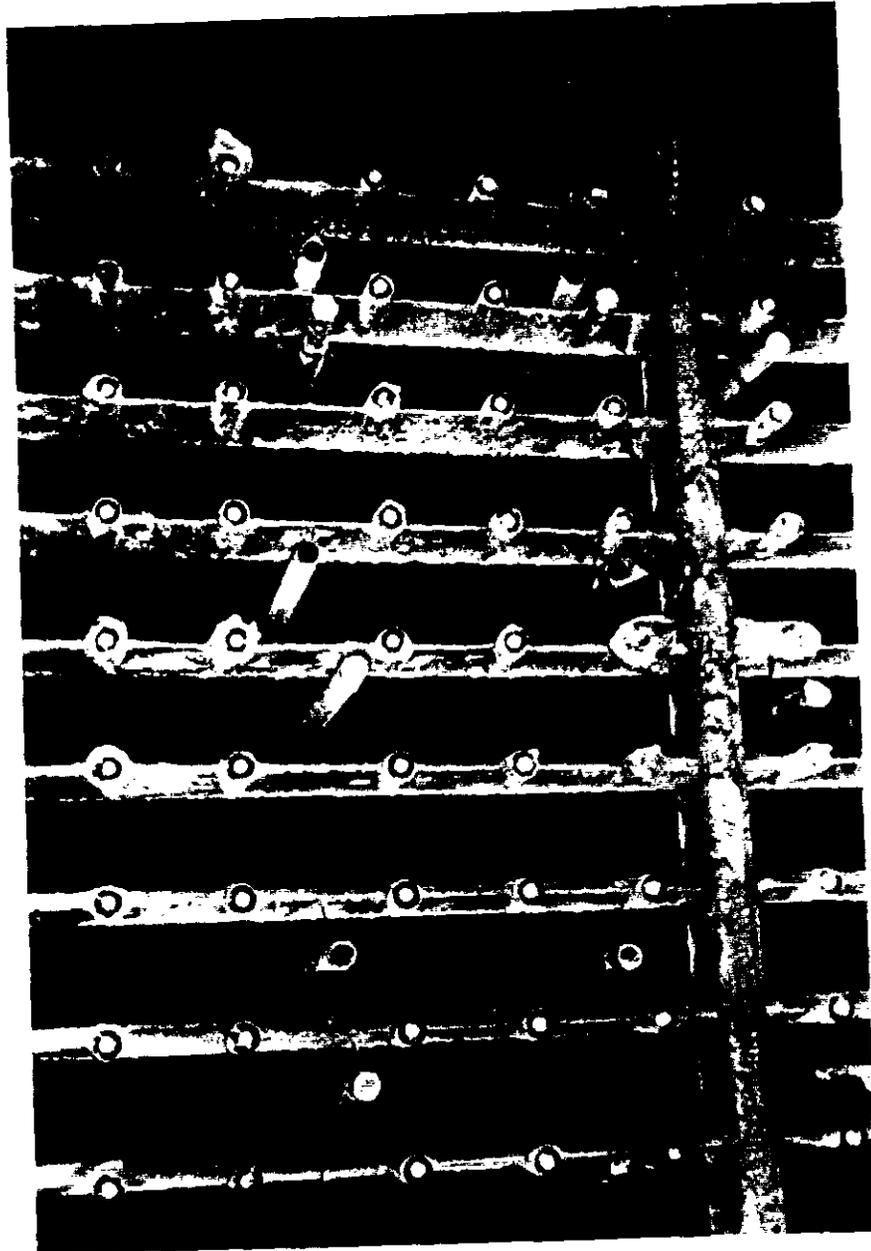


Figure 28. Front view of water sprays showing no buildup of solids on lances after 11 continuous days of operation.

atomizers located around the injector discharge. The bottom row of injectors were taken out of service because the injected solids quickly deposited on the humidifier floor. Also, with nine injectors in service, the injection air velocity was low and injector plugging occurred. Even with only five of the nine injector pipes in operation, the distribution of hydrated lime solids in the flue gases appeared to be adequate.

When recycle solids were fed, normally eight of the nine recycle solids injectors were operated. As with the hydrated lime, the center injector was taken out of service to prevent solids buildup on the atomizers around the center recycle injector. Unlike the hydrated lime injectors, the bottom three injectors of the recycle solids injector pipe array were placed in service. These bottom injector pipes remained open because the conveying air rate was about three times that available for the hydrated lime injection.

## HUMIDIFIER CHAMBER

### Flue Gas Velocity Profiles

Humidifier flue gas velocity profiles were measured during the Coolside process equipment checkout phase of operations. This was done to determine if additional flue gas flow straightening would be required. Pilot-scale tests<sup>11</sup> demonstrated that establishing a symmetric and flat inlet gas velocity profile was desirable to prevent water spray impingement on the duct walls.

Figure 29 shows the general layout of the humidification chamber. Flue gas from the air heater made three horizontal right angle bends before entering an expansion section at the inlet of the humidifier. At the humidifier inlet, the flue gas passed through an array of air foils which contain one hundred 0.8 GPM atomizer nozzles. The atomizer nozzles produce a water spray which completely evaporates by the time the flue gas exits the humidifier. From the humidifier, the flue gas is conveyed to the ESP for particulate removal.

The humidification chamber was equipped with three sampling locations, each containing seven ports (see Figure 29). Velocity measurements at these ports were made with a calibrated "S" type pitot tube at three operating conditions:

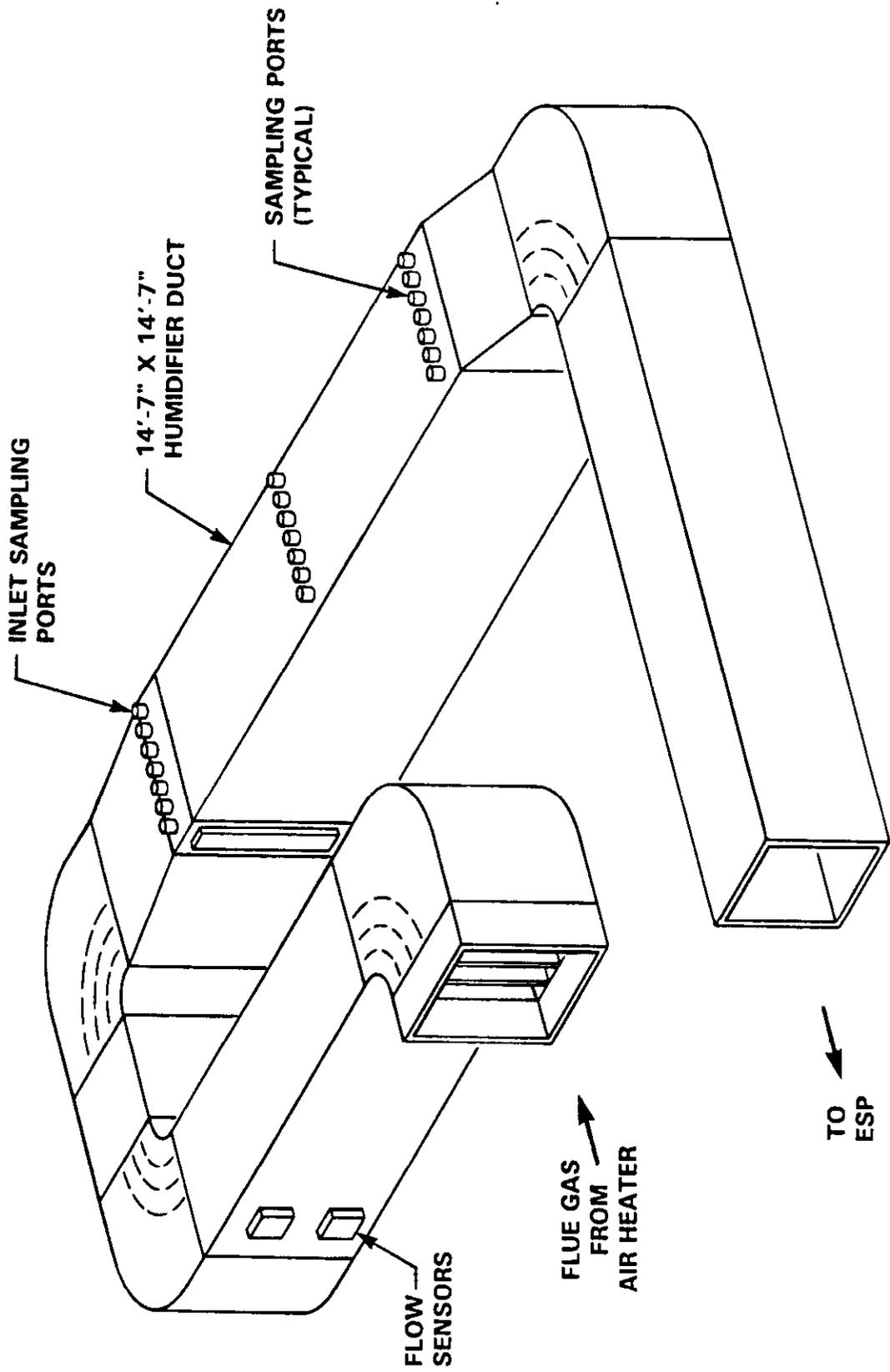


Figure 29. General arrangement of humidifier duct work.

- Case 1            High flue gas flow with atomizing air off  
                  (Figures 30a, 30b and 30c)
- Case 2            Low flue gas flow with atomizing air off (Figures  
                  31a and 31b)
- Case 3            Intermediate flue gas flow rate with atomizing air on  
                  (Figures 32a and 32b).

No attempt was made to measure velocity profiles with water to the sprays because accurate pitot tube differential pressure readings cannot be obtained in a hot gas stream with large amounts of water droplets and moist sticky solids.

The velocity data for the three cases (Appendix B, Table B-1) are presented in Figures 30a to 32b as surface and contour plots. Measurements were taken in the 14'-7" square humidification chamber duct on 49 point grids except for Test 3 (Figure 30c) which used a 21 point grid.

Figures 30a, 31a and 32a clearly show that at the humidifier inlet, the velocity along the south wall was substantially higher than along the north wall. The two close-coupled, right angle 90° turns immediately upstream of the humidifier were likely responsible for forcing gas flow towards the south wall. In the vertical direction, the inlet velocities tended to be rather uniform. As one would expect without atomizing air on, the gas velocity profile (see Figures 30b, 30c, 31b and 32b) tended to equalize and flatten out as the flow moved towards the humidifier outlet.

For the intermediate flue gas flow case (Case 3), the atomizing air was turned on to determine the effect of the high pressure air momentum on the flow patterns. The aspirating effect of the atomizing air appeared to slightly skew the gas flow at the humidifier inlet by further increasing the flow along the south wall. This is seen from a comparison of the ratio of the south side to north side average inlet gas velocities. For Case 1, the average inlet flue gas velocities at 1.3 ft and 13.3 ft from the north wall were 20.6 fps and 34.1 fps, respectively, and the ratio is 1.66. For Case 2, the respective average

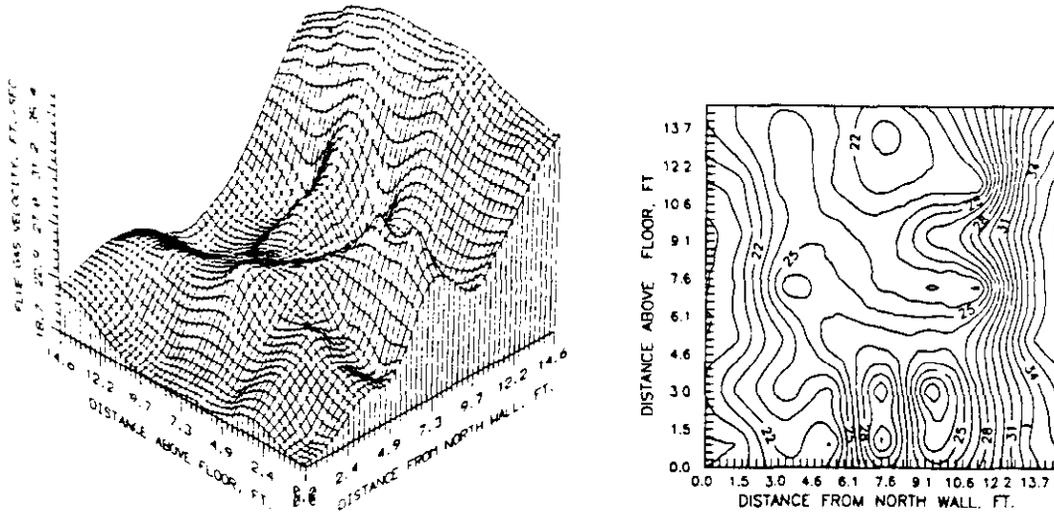


Figure 30a. Humidifier inlet velocities (high flow case -- atomizing air off).

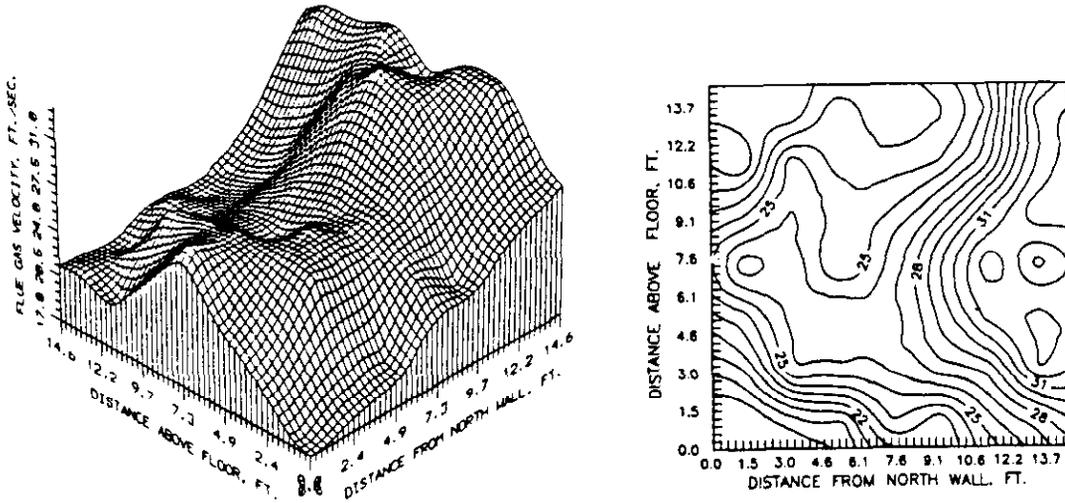


Figure 30b. Humidifier center velocities (high flow case -- atomizing air off).

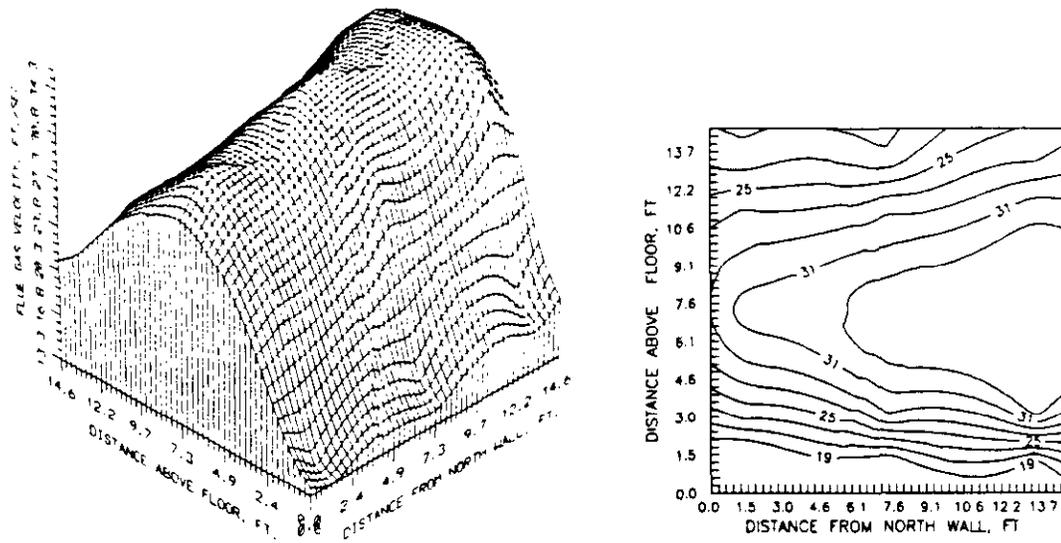


Figure 30c. Humidifier outlet velocities (high flow case -- atomizing air off).

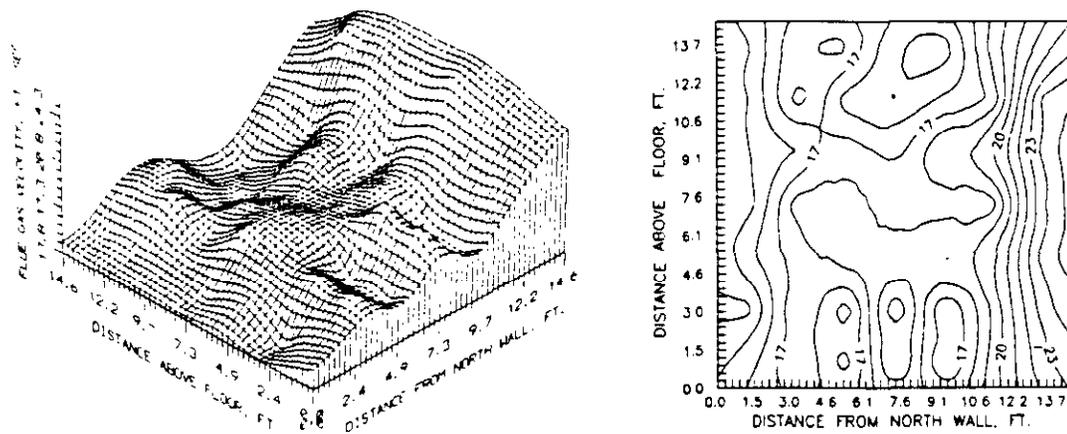


Figure 31a. Humidifier inlet velocities (low flow case -- atomizing air off).

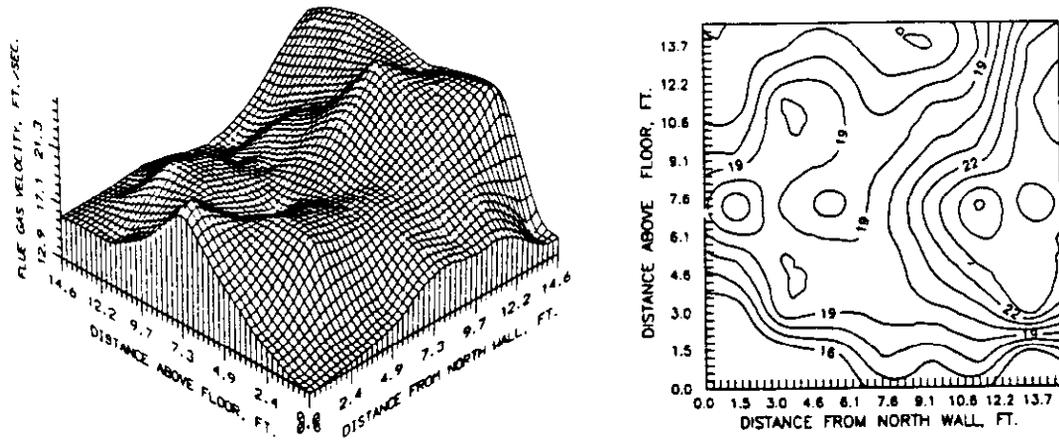


Figure 31b. Humidifier center velocities (low flow case -- atomizing air off).

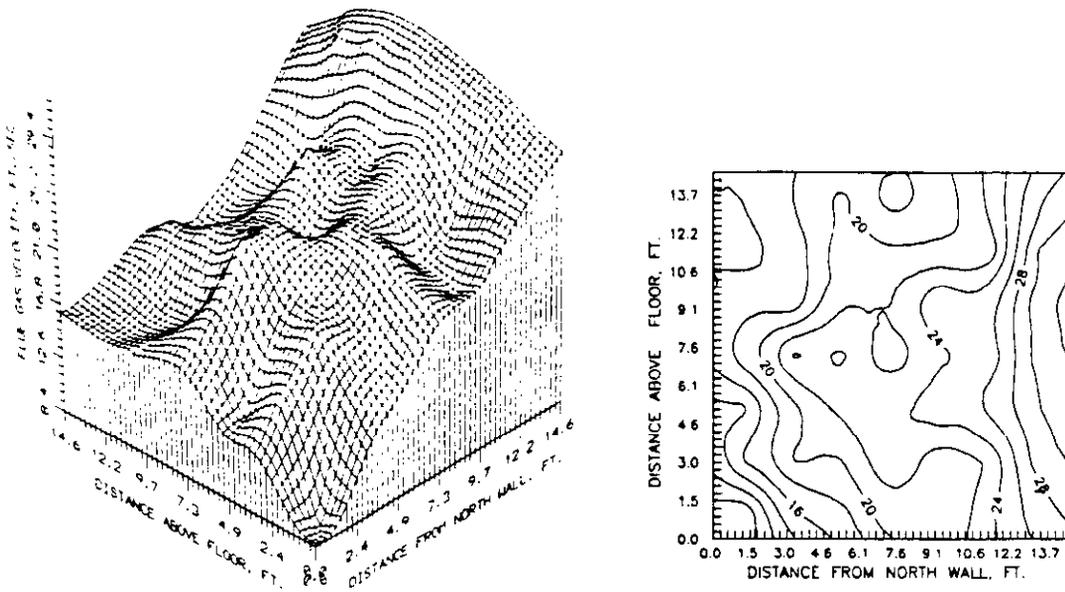


Figure 32a. Humidifier inlet velocities (intermediate flow case -- atomizing air on).

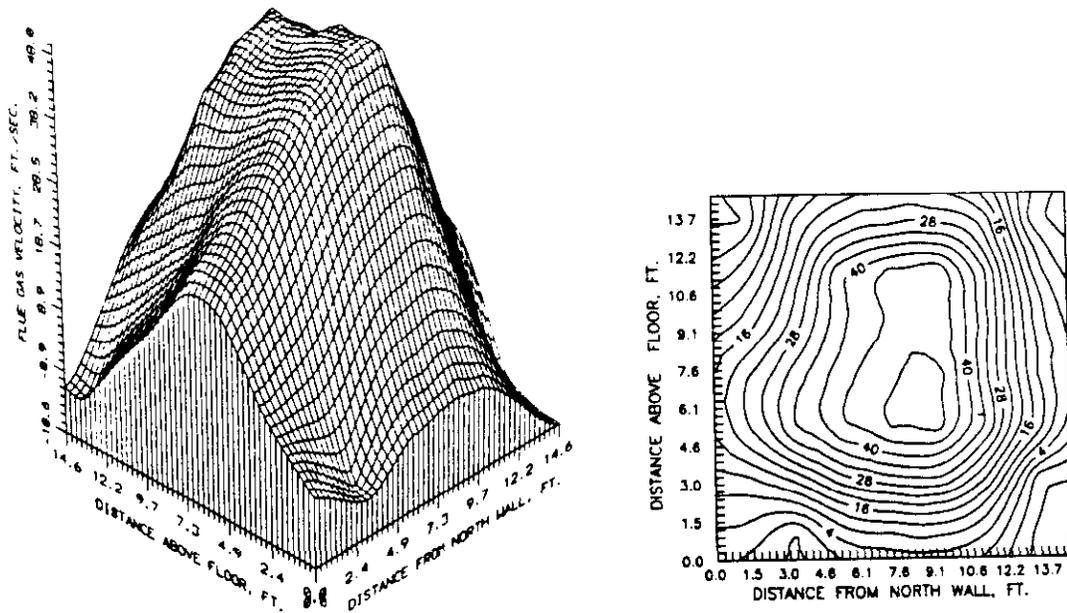


Figure 32b. Humidifier center velocities (intermediate flow case -- atomizing air on).

velocities were 14.9 fps and 23.5 fps with a ratio of 1.58. The ratios are nearly identical for the two cases. However, for Case 3, the velocity ratio increased significantly to 1.94 with respective velocities of 15.2 fps and 29.4 fps.

A more important effect of the atomizing air was the channeling and acceleration of flue gas flow towards the center of the humidifier. This is clearly indicated by the velocity profile and bull's-eye contour pattern shown in Figure 32b. So much flow was forced toward the center that negative or reverse flow patterns developed in the corners and on the floor of the duct. From the standpoint of solids dropout and removal, the low gas velocities along the floor are not desirable. Methods to accommodate this condition are discussed under "Chamber Deposits" in this section.

Ideally, a flatter velocity profile is desirable at the inlet with velocity deviations randomly distributed and all velocities within  $\pm 10\%$  of the average. This could have been done by redesign of the humidifier inlet turning vanes for

flow distribution. However, because of the time and expense involved with modifying the turning vanes and because of the relatively concentric flow profile which was established downstream of the atomizers due to the use of the atomizing air, the modification was not felt to be necessary for the Edgewater test program. In retrospect, redesign of the inlet turning vane likely would have improved the humidifier operation somewhat by eliminating the flow bias toward the south side. This often required reducing the water flow to the north half of the atomizer lances to obtain a more uniform outlet temperature profile (see this section; "Water Spray Lance Design And Operation").

### Inlet Flue Gas Flow Measurement

Continuous measurement of the flue gas flow rate to the humidifier was an integral part of the Coolside process control equipment. The locations of the thermal dispersion mass flow meter sensors are shown in Figure 29. An accurate measurement of the flue gas rate to the humidifier is desired to provide a backup means of checking calculated rates based on boiler operating conditions and flue gas compositions.

The pitot tube tests described in the previous section were used to confirm the accuracy of the flow instrumentation. The on-line flow unit consisted of two probes each having three sensing heads. The mass flow rate of the flue gas was determined from the temperature difference between matched pairs of heated and unheated platinum resistance temperature detectors (RTDs). The temperature difference was greatest under no-flow conditions.

Because of plant computer data logger problems, the on-line flue gas flow rate was not recorded when Case 1 pitot tube velocity measurements were taken. The following measurements were however obtained for Cases 2 and 3.

**Table 11**  
**COMPARISON OF MEASURED HUMIDIFIER INLET**  
**FLUE GAS FLOW RATES, lb/hr**

	On-Line Measurement	Based On Pitot	Difference %
Case 2	720,000	718,120	+0.26
Case 3	850,000	864,620	-1.69

As shown in Table 11, the on-line instrument results compare very well with the pitot tube measurements. The expected accuracy of this particular instrument configuration was 3.5 % of full-scale (1.5 MM lb/hr) or  $\pm 52,500$  lb/hr.

### Water Spray Lance Design And Operation

#### General Description--

A sketch of the atomizer array layout is shown in Figure 33. The design was based on the results of quarter scale humidifier testing conducted by B&W<sup>12</sup>. The same atomizer array was used for the EPA LIMB tests which preceded the Coolside program. The array consisted of one hundred 0.8 gpm B&W I-Jet atomizers. The atomizers used compressed air, about 120 psig, to produce a very fine water spray mist. Rather than use single 14'7" lances which would totally traverse the humidifier cross section, a split lance (7'3.5" long) design with a center support was provided. This reduced the weight by half and made lance installation and removal easier. There were twenty horizontal atomization lances (ten per side) each containing five atomizers. Originally, the atomizer array consisted of 110 nozzles on 22 lances. Prior to the Coolside testing, the bottom row of atomizers was removed to provide additional clearance between the spray plume and the floor.

Atomizers were spaced sufficiently far apart to minimize nozzle-nozzle interactions which can lead to droplet coalescence. The original lance design provided several different atomizer spacing options which could be adjusted in the field. The non-uniform spacing shown in Figure 33 was not done for process reasons, but rather was the result of piping limitations caused by the atomizer hookup arrangement options provided by the original design.

The major equipment and control instrumentation for the water atomizers is shown in Figure 34. For clarity, the piping details, which are typical for all lances, are shown for only one lance. Additive-containing water is pumped from the water storage tank to the atomizer lances. The water rate is automatically controlled by the humidifier outlet temperature controller through a cascade control system which resets the set point (SP) of water flow controller. Air is supplied to the atomizing lances through a pressure controller. When necessary,

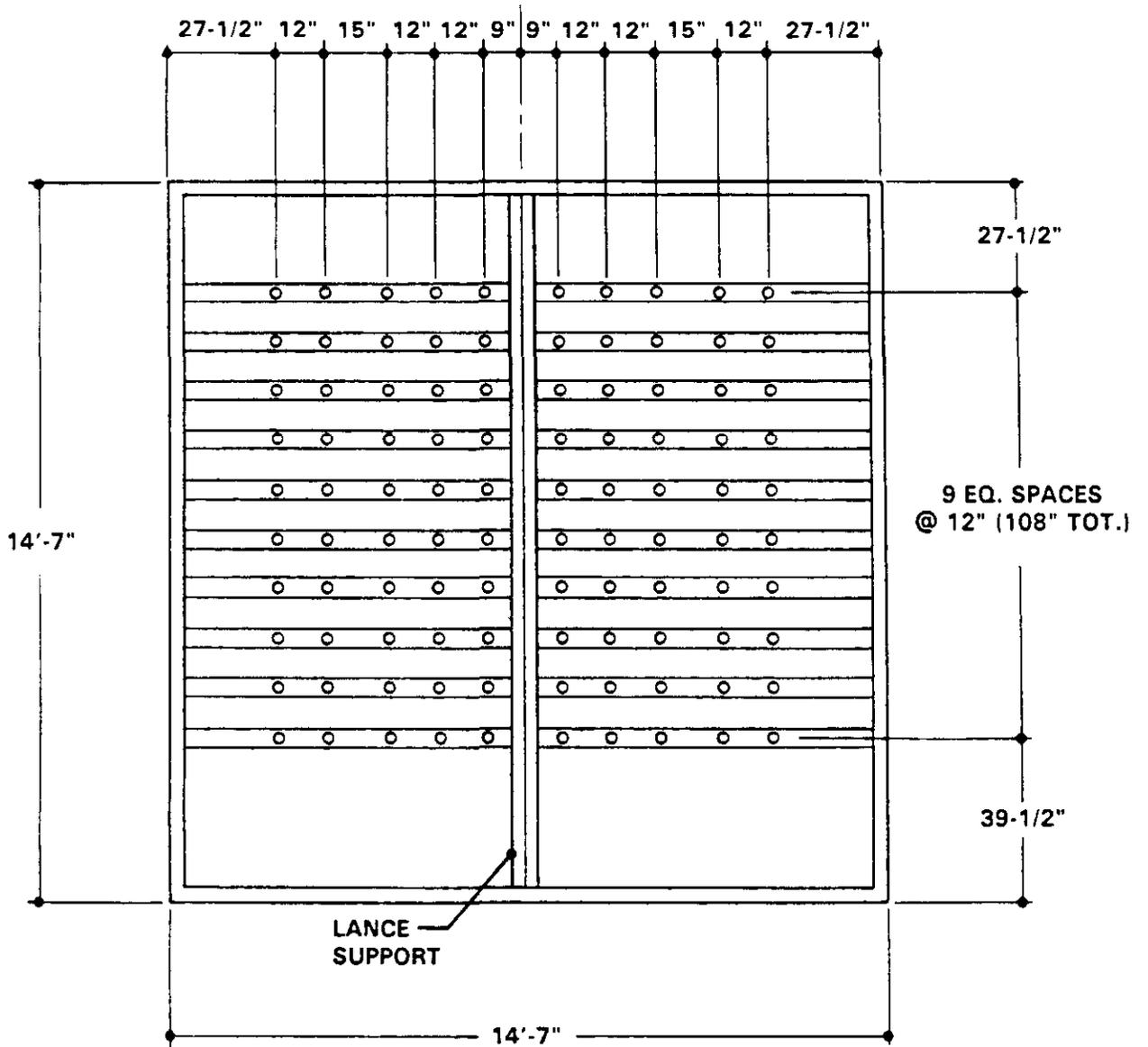


Figure 33. Edgewater Station humidifier atomizer array layout.

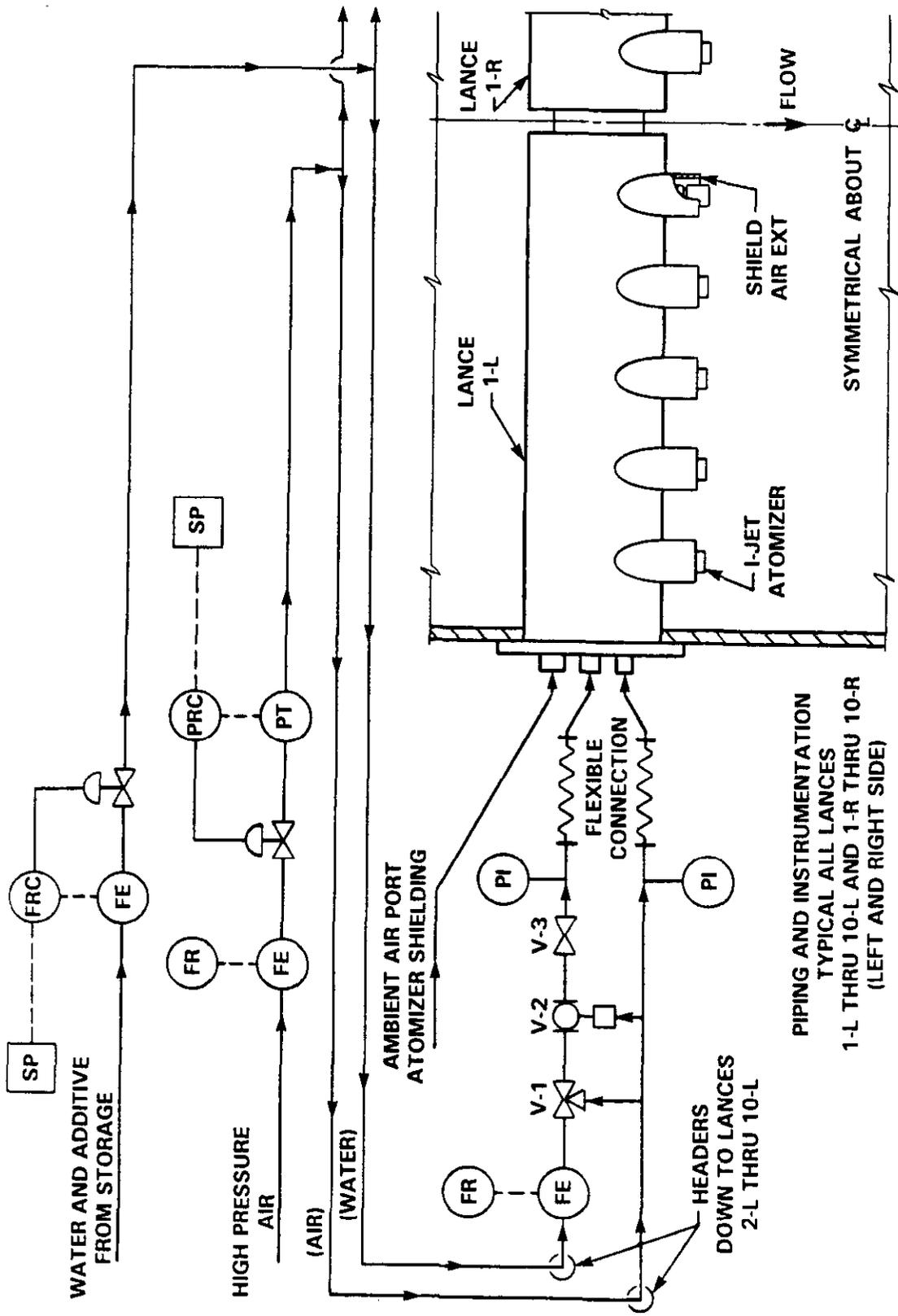


Figure 34. Edgewater Station process control system for atomization lance array.

the air rate is manually adjusted to maintain a minimum air-to-liquid ratio by changing the pressure controller set point.

To enable detection of lance operating problems such as atomizer plugging, each water feed line was equipped with a flow element (FE--see Figure 34). Magnetic flowmeters are used for this application because of their high accuracy, high turndown, and low pressure drop features.

Other features of the lance hookup include: (1) the use of a manual three-way valve (V-1) between the high pressure air and the water supply piping to allow rapid and complete purge out of the water line prior to shutdown, (2) the use of automatic water shutoff valves (V-2) which were activated by loss of atomization air pressure, and (3) the use of manually adjusted globe valves (V-3) to trim or balance the water flows between lances. To facilitate lance removal and repairs, flexible hose connections with quick disconnects were used to attach the air and water supply lines.

The manual globe valves were used to redistribute the humidification water between the north and south side humidifier water spray lances and between the top and bottom spray lances to reduce temperature gradients at the humidifier outlet.

The air and water supply system worked well. Only two minor problems were experienced during the Coolside testing; cracking of the rubber high pressure air hoses due to the high operating temperature (190°F) and external corrosion of the aluminum actuators on the loss-of-air pressure, automatic water shutoff valves. To eliminate the hose deterioration problem, the rubber hoses were replaced with braided sheath all-steel hoses.

The corrosion of the automatic water shutoff valve actuators was due to contact with small amounts of caustic solution from leaks (drips) around screwed stainless steel pipe connections. The caustic water leaks were easily repaired by replacing leaking stainless fittings with galvanized steel pipe nipples. However, because many of the shutoff valve actuators had been severely damaged, the valves were all removed and not replaced near the end of the Coolside testing program. For any commercial humidifier design, an automatic water shutoff system

is still recommended on the water headers to the spray lances. The use of stainless or carbon steel valve actuators would eliminate the potential for corrosion from caustic solutions.

#### Droplet Size Considerations--

Humidifier design requires a knowledge of atomizer performance as a function of capacity and air/liquid ratio. Since the largest droplets take the most time to evaporate, the humidifier residence time is determined by the weight fraction and size of the largest droplets produced by the atomizing nozzles. As an example, using the Marshall correlation for evaporation of pure liquid droplets<sup>13</sup>, the evaporation times for 10 $\mu$ m, 50 $\mu$ m, and 100 $\mu$ m water droplets were estimated: 0.026, 0.66, and 2.6 seconds, respectively, assuming a 300°F inlet flue gas temperature, 125°F adiabatic saturation temperature and a 20°F approach to the adiabatic saturation temperature. Assuming an atomizer produces a maximum droplet of 100 $\mu$ m, the calculation indicates that the humidifier should be designed for a minimum of 2.6 seconds residence time to insure complete evaporation within the humidifier. The Edgewater humidifier design provides about 2.4 seconds residence time based on a projected full boiler load flue gas rate of 1,037,000 lb/hr. In practice, to sustain operation at a 20°F approach to the adiabatic saturation temperature, the gas flow to the humidifier had to be limited to a maximum of 850,000 lb/hr or less (see Humidifier Control section).

The Edgewater humidifier design (14'7"x14'7"x 56' [center line of water spray lances to outlet]) was the result of pilot-scale "Operability Drying Time Tests" using commercially available atomizers in a 3'x3'x40' duct; quarter scale "System Design Validation Tests" in a 6'x6'x80' duct; and atomizer development efforts<sup>12,14</sup>. Atomizer performance tests conducted in conjunction with the 3'x3'x40' evaporation tests clearly established that there were significant differences in the performance of two-fluid atomizers which were operated under similar conditions.

A comparison of atomizer performance of four dissimilar-design, small-capacity, commercially available atomizers and two B&W atomizers is shown in Figure 35. In the figure, the Sauter mean diameter is plotted against air/liquid ratio. The water flow rate range (gallons/minute, gpm) and atomizing air

pressure range (lb/in<sup>2</sup>, psi) for each atomizer are also shown. The results show that for these atomizers, there were two distinct performance groupings. The two B&W atomizers were in the grouping which performed the best and produced the finest droplet diameters at a given air/liquid ratio. Because of this and because the B&W atomizers had a rugged design and could be readily modified to optimize performance, the B&W I-Jet atomizer design was selected for the quarter-scale and full-scale tests.

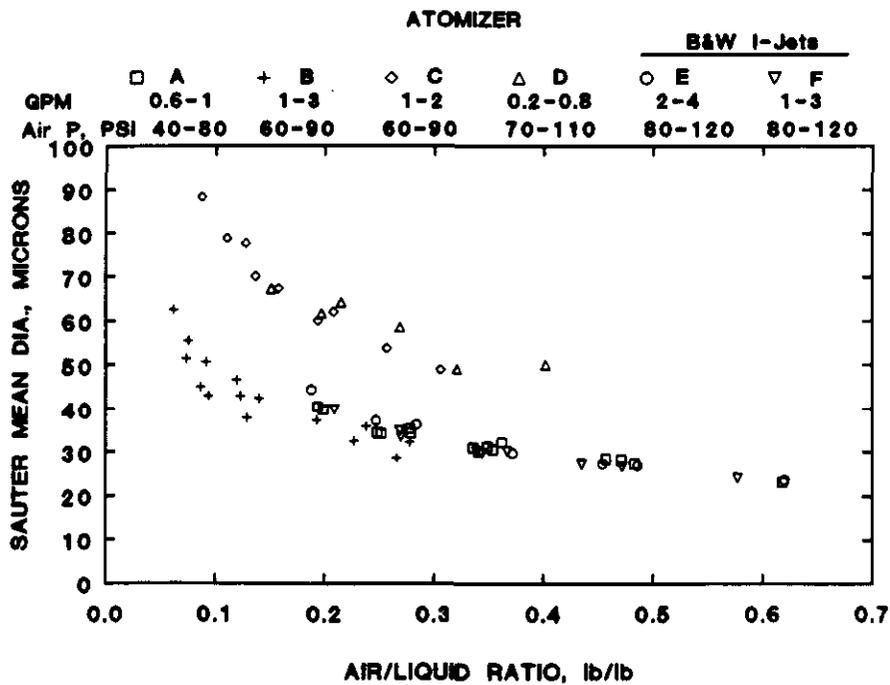


Figure 35. Performance of several atomizers.

#### Atomizer Deposits--

Because the water spray atomizers operate in an extremely dusty environment, special precautions were required to prevent deposit formation at wet/dry interface areas on the atomizers. Atomizer deposits will interfere with atomizer performance and can cause atomizer dripping and spray coalescence. The presence of lime-containing solids in the flue gas posed special problems since these solids can, under moist conditions, readily react with SO<sub>2</sub>, SO<sub>3</sub>, CO<sub>2</sub>, and fly ash to form cementitious deposits.

To minimize atomizer deposit formation, the water spray lances were specially designed to have a streamline airfoil shape for reduced turbulence. Additionally, the airfoils were hollow and had annulus tubes around each atomizer through which ambient air (shield air) was drawn by the pressure differential between the outside and inside of the humidifier. The shield air acted as a buffer between the atomizers and the dusty flue gas to prevent solids contact with the atomizers.

Several shield air designs were tested during the "Operability Drying Time Tests." The tests were conducted in a 3'x3'x40' humidification test section using flue gas from a coal-fired Stirling boiler at B&W's Alliance Ohio Research Facility<sup>14</sup>. During these short-term (less than eight hour) tests, the B&W airfoil shield design appeared to perform the best.

The water atomizer lances originally installed in the Edgewater humidifier are pictured in Figure 36. The shield air annuli around the atomizers are on the trailing edges of the spray lances.

The flue gas from the Stirling boiler contained coal fly ash but did not contain lime sorbent materials ( $\text{Ca}(\text{OH})_2$  or  $\text{CaO}$ ). At Edgewater where the flue gas contained both fly ash and sorbent materials, severe atomizer deposit formation was experienced during early operations. Figure 37 shows typical tulip-shaped water lance deposits which formed quickly during initial shakedown operations, often in a few hours when the humidifier was operated at a close approach to saturation. Frequently, the deposits also covered the shield air annuli. To control the lance deposits, two changes were made by B&W. First, a drop hammer rapper system was installed external to the humidifier. The hammers, which are used in some ESPs to rap plates and wires, were used to continuously rap the water lance support structures. *This did not prove effective in keeping the lances clean.*

The second change, which did prove to be effective, was to redesign the airfoil. B&W had the atomizer lance flow modeled in a wind tunnel. These tests led to a redesign of the airfoil. The airfoil shape and the size and length of the annulus tubes (nacelles) around the atomizers were modified. The redesigned

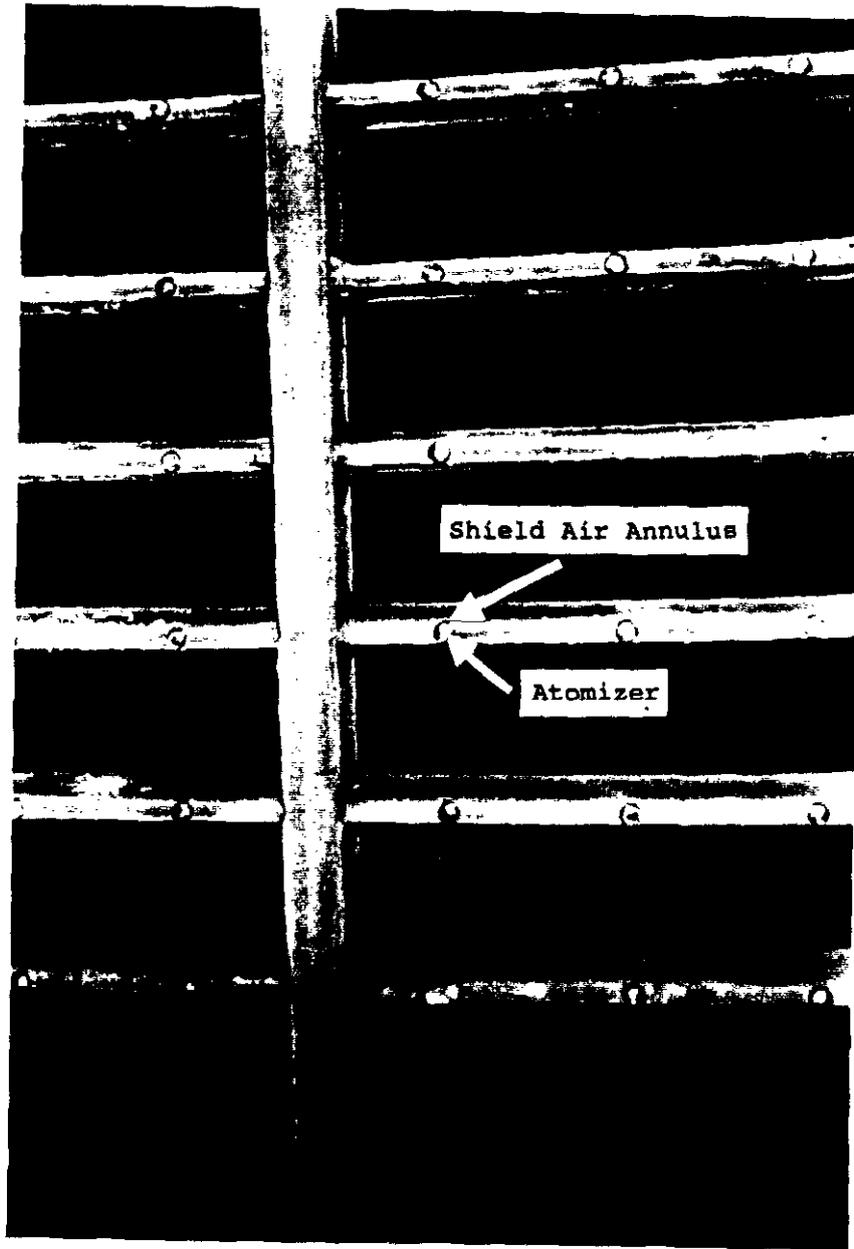


Figure 36. Original atomizer lance design.

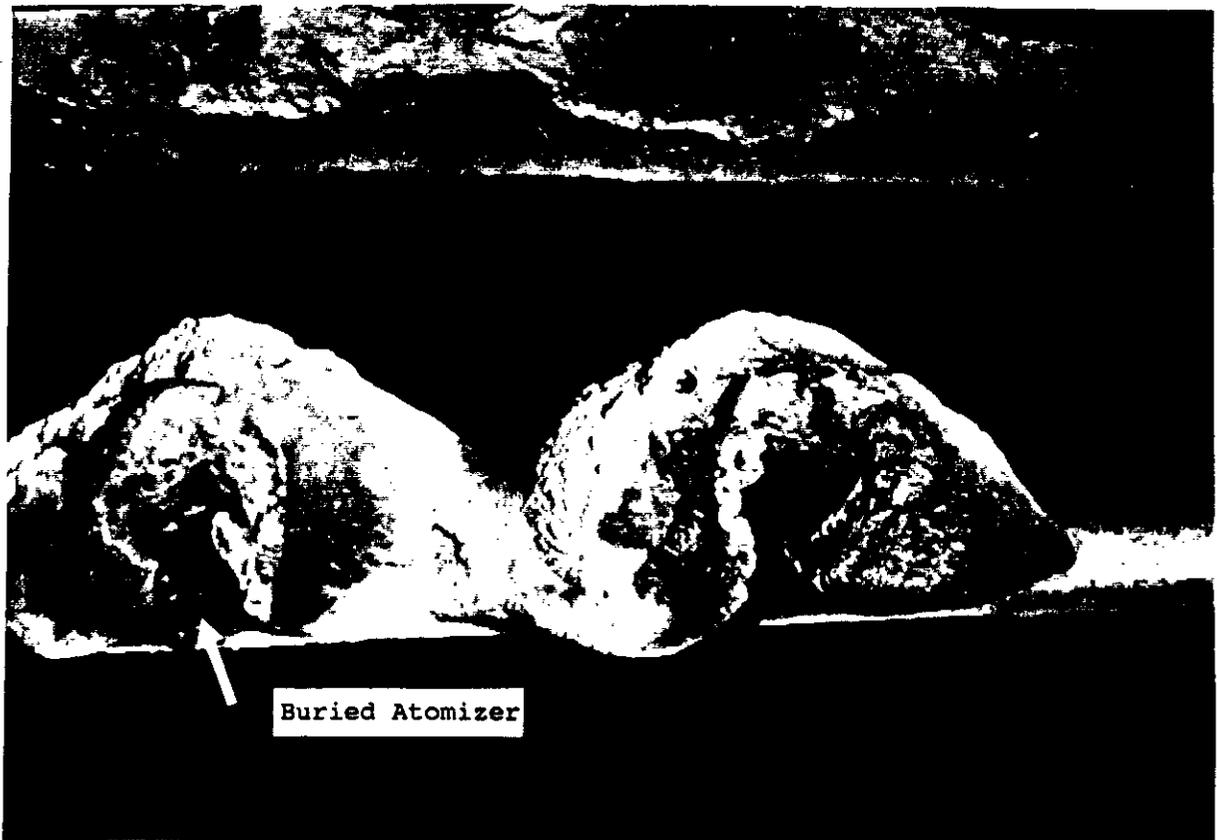


Figure 37. Atomizer deposits (tulips).

lances are shown in Figure 38 after an 11-day run during which the humidifier was operated round-the-clock at a 20°F approach to the adiabatic saturation temperature. As shown, most of the atomizers were free of large deposits.

A completely passive deposit control design probably will not be adequate for commercial applications of the technology since the humidifier will be on-line for many months. The design will likely need to be augmented with soot blowers or a mechanical cleaning system to remove the occasional lance deposit.

#### Humidifier Performance Degradation--

When new and at design operating conditions, the B&W I-Jet atomizers produced almost no droplets above about 110  $\mu\text{m}$ . This fine spray feels cool to

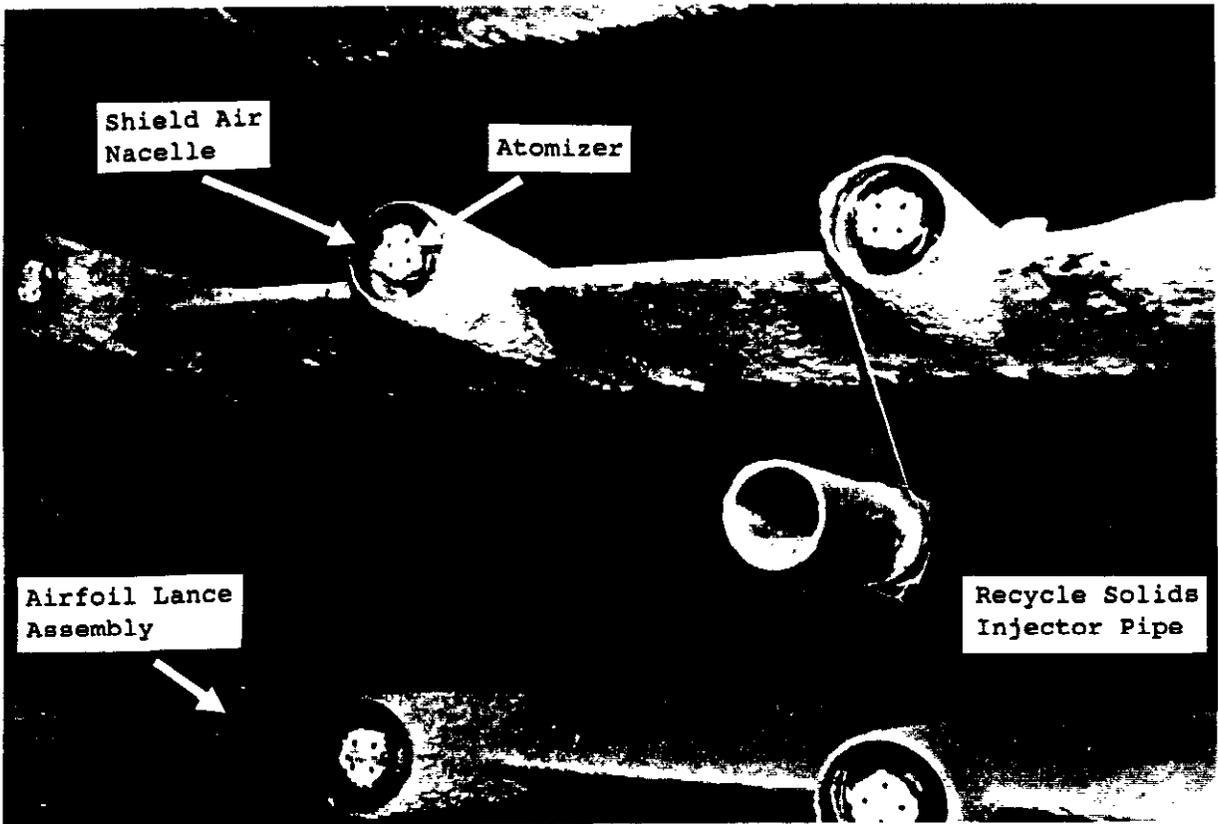


Figure 38. Closeup view of improved atomizer lances after 11 days of operation.

the touch of a hand placed in the spray two to three feet downstream of the atomizer. Individual droplets (stingers) cannot be felt. This simple, but effective, method of checking the spray fineness is referred to as a wrist test. The wrist test was used to identify gross defects in newly assembled spray lances before installation in the humidifier and to check the in-place lance performance during subsequent operations.

After several months of Coolside process operations, a wrist test was performed. The test showed that over 10% of the atomizers were producing a coarser than expected spray. Close external inspection of the atomizers showed that many were beginning to develop a teardrop-shaped exit port erosion pattern

like that shown in Figure 39. The atomizer erosion was likely caused by either high velocity water droplet abrasion or to the presence of fine silt in the lake water used for humidification.

The atomizers used for the Coolside testing were not hardened and were made of 316 SS. The erosion problem can likely be corrected by hardening the atomizers via metallurgy change or the use of ceramic inserts such as those used in spray dryer applications.

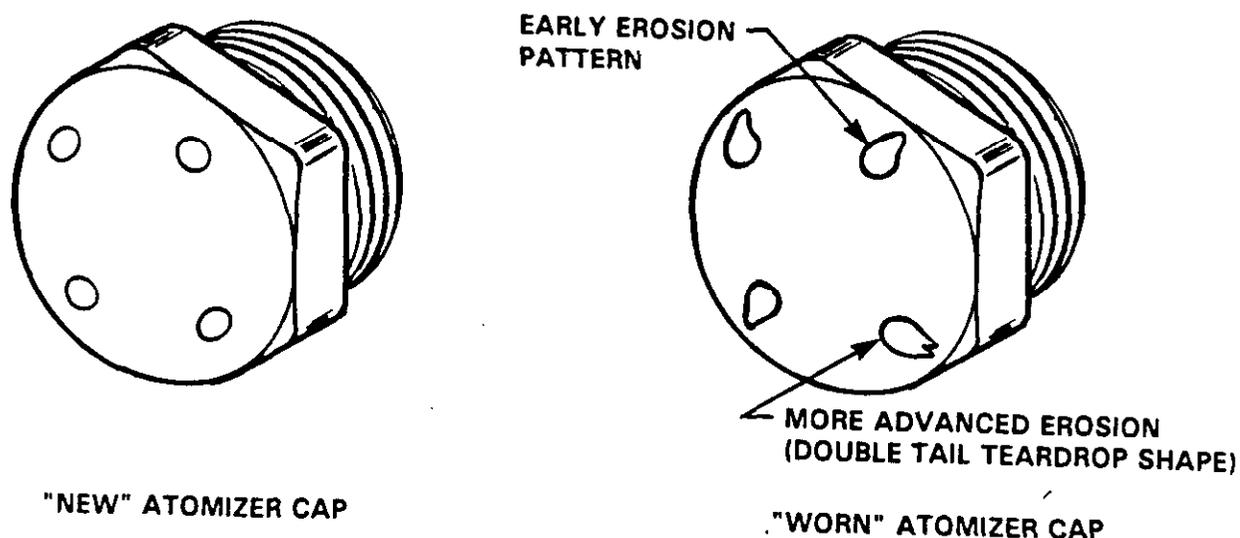


Figure 39. I-Jet atomizer discharge port erosion.

When the atomizer erosion was detected, it was assumed to be responsible for a change in droplet size distribution. There was insufficient time, however, to confirm this and to purchase and install new atomizers. To regain atomizer performance, the air/liquid ratios were maintained as high as possible by increasing the air pressure at the atomizers and by limiting the gas flow through the humidifier which limited the required water rate. Additionally, the humidifier outlet temperature was increased to provide a 25°F rather than a 20°F approach to the adiabatic saturation temperature. These changes improved the

humidifier operation by helping to insure complete droplet evaporation within the chamber. Subsequently, the degradation of atomizer performance was found to be due to plugging of the air holes by dirt, scale, and a tarry material (probably oil carryover from the air compressor). The most affected atomizers were located at the end of the lance furthest from the air and water supply line inlets.

### Chamber Deposits

#### Solids Dropout--

Unless flue gas velocities are maintained high, solids dropout can become a problem in horizontal humidifier designs. At Edgewater, due to site constraints, a horizontal humidifier design was selected for installation on top of the boiler house roof. Space limited the maximum straight length to about 56 feet from the centerline of the atomizer array to the outlet. Because a minimum residence time of 2.6 seconds was required for evaporation, the humidifier full load flue gas velocity was about 22 fps. Additionally (see Figure 32b), flue gas velocities just above the floor were near zero with atomizers operating. Under such conditions, solids dropout became a problem as can be seen in Figure 40 where the solids layer on the floor is two to three feet deep.

The floor solids were normally dry unless the accumulations built up to intersect the water spray plume. The problem was nearly eliminated by installing thirteen 2" diameter dust blower pipes at 4' to 5' spacings along the floor as shown in Figure 41. Each pipe had fourteen 1/8" diameter holes spaced one foot apart. The pipes were installed so that air flow from the holes blew parallel to the floor toward the humidifier outlet. This helped to fluidize the floor solids and push the solids toward the outlet where the duct size decreases from 14'-7" x 14'-7" to 10'-3" x 10'-3". The reduction in duct cross section at the outlet increased the flue gas velocity enough to carry the solids to the ESP where the solids were removed.



Figure 40. Humidifier floor solids accumulations (early operations).



Figure 41. Dust blower pipe installation in humidifier.

The blower pipes were operated individually and sequentially. The operation was continuous with each blower pipe pressurized for a predetermined time, normally 30 seconds, before the air supply was switched to a different blower pipe. The nominal air consumption was about 900 to 1200 lb/hr at air supply pressures of 10 to 15 psi, respectively. Pressure inside the dust blower pipe was initially set by adjusting a hand globe valve ahead of the air supply solenoid valve. Figure 42 shows the control details.

Figure 43 shows the condition of the humidifier after the last Coolside process run which lasted eleven days. The view is looking toward the humidifier outlet. During this run, the humidifier operated round-the-clock at a 20-25°F approach to the adiabatic saturation temperature. The view shows only small

6"-8" high piles of floor solids which collected on the back sides (side opposite air holes) of the blower pipes. Between the blower pipes, the floor was bare or covered with a thin layer of dust.

The floor dust blower system used very little high pressure air and worked well for the relatively short Coolside process operating periods. For a commercial system, a more vigorous blower system will be required to insure removal of coarser solids debris from wall scales which can build up over time.

#### Wall Scales--

Figure 43 shows the presence of wall scales; some of which have spalled or sloughed off the walls during humidifier shutdown. These scales normally grew to a thickness of two to four inches and then begin to slough. The scales, which were quite soft, apparently form when solids which contain small amounts of moisture impact the walls, stick, and then dry out. Similar scales often form in spray dryer systems.

No special efforts were taken to remove the deposits during operations. Normally, the wall scales were not removed between runs. However, during a shutdown, the scales could be rapidly removed by simply hitting the base of the inside walls with a sledge hammer. When done, personnel were in a pressurized air suit because of the extreme dust conditions which immediately resulted from solids falling off the walls. For a commercial system, external mechanical wall rappers or possibly a sonic horn system could be installed to minimize the thickness of the wall scales. This would help minimize the difficulty of transporting clumped spalled material which would fall to the floor of horizontally designed humidifiers. Additionally, for commercial humidifier designs whether vertical or horizontal, provision should be made for an outlet solids removal hopper. This would allow positive removal of scale deposit materials which would fall into the hopper for a vertical humidifier design or which would be blown or pushed to the hopper by floor soot blowers if the humidifier was a horizontal design. The humidifier hoppers should be designed to handle material with poor flow characteristics.

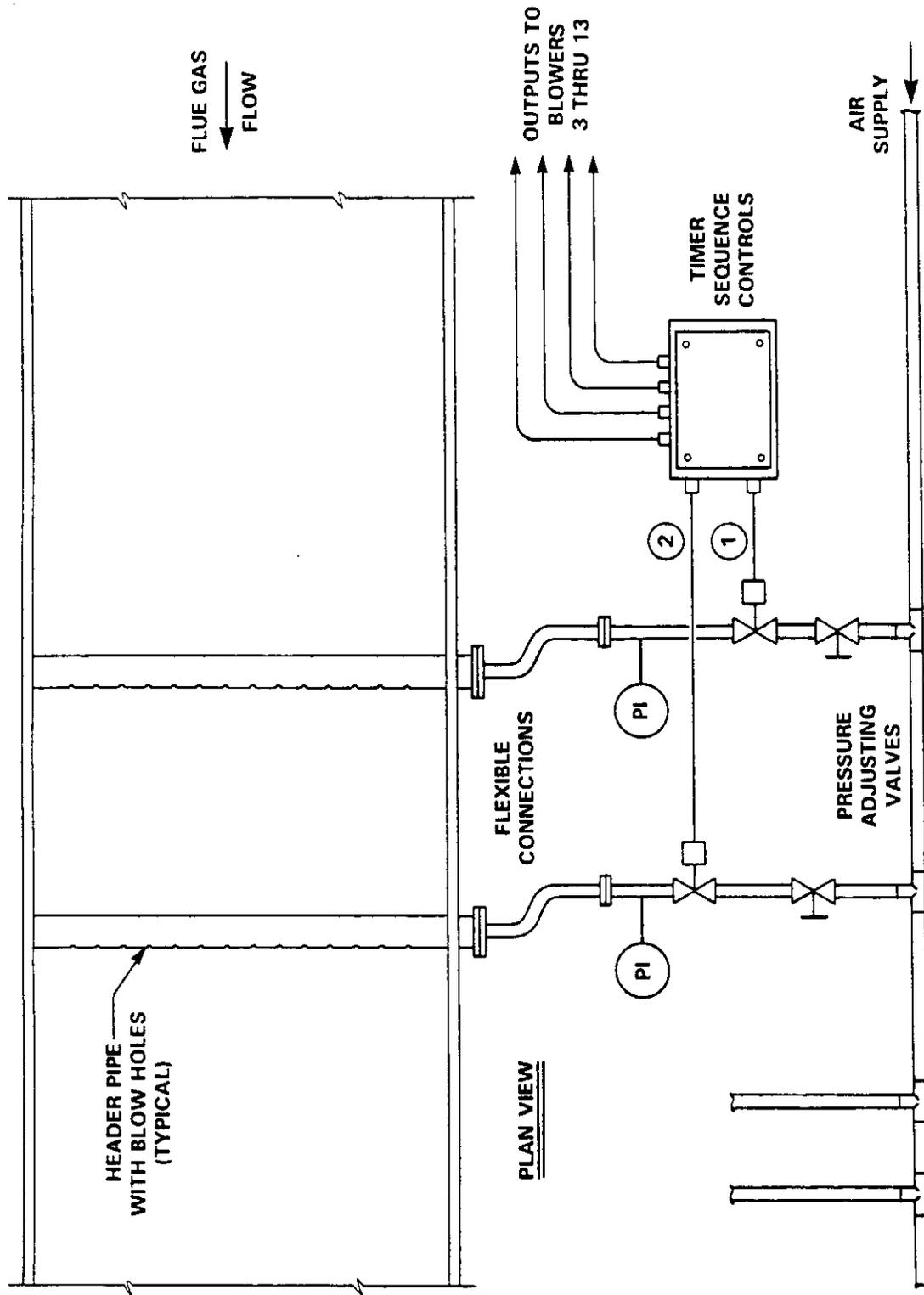


Figure 42. Humidifier floor dust blower control system.

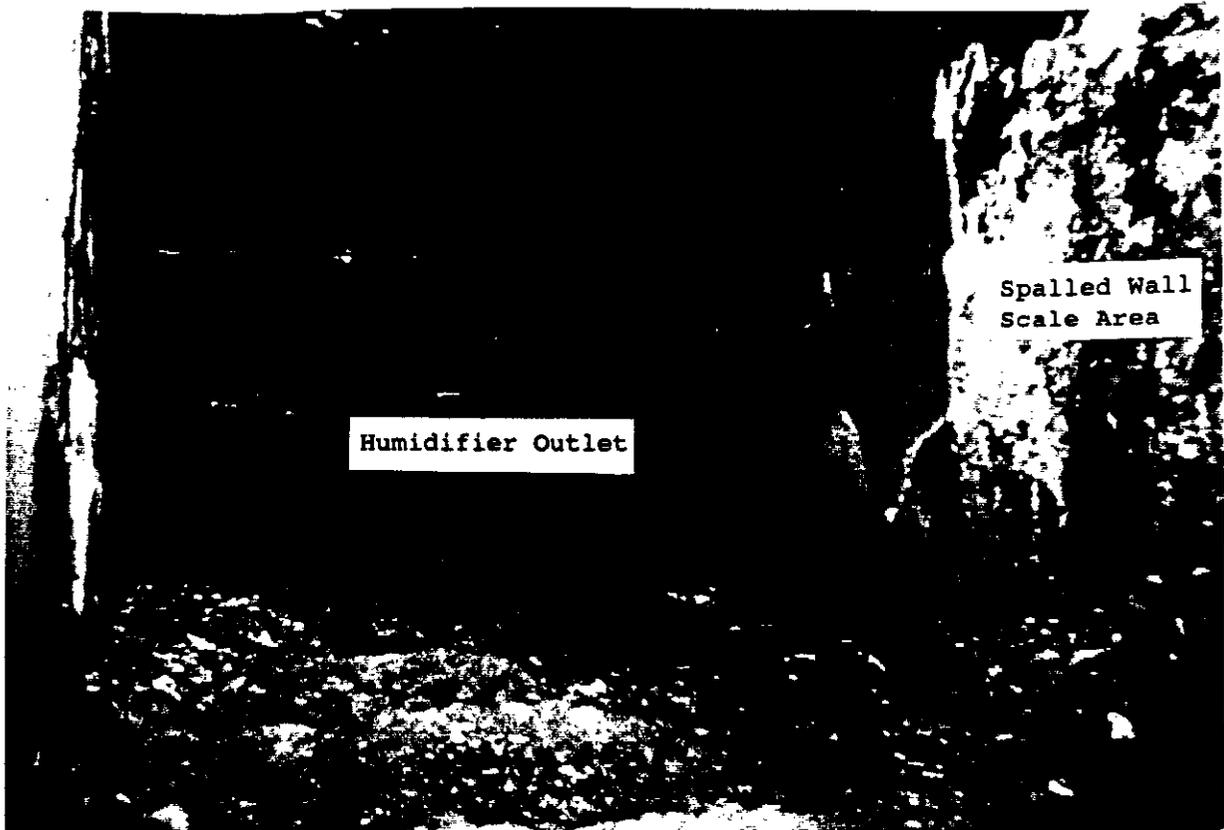


Figure 43. Humidifier floor solids accumulations after installation of floor dust blower system.

#### Outlet Turning Vane Deposits--

Turning vanes were provided at both the inlet and outlet of the humidifier. Because the dust material at the inlet is completely dry, no deposits formed on the inlet vanes. However, scales nominally three to six inches thick formed on

the front faces of the outlet turning vanes as shown in Figure 44. These scales are harder than the humidifier wall scales because the change in flow direction and the higher velocity of the flue gases at the outlet increases the force of wall impact of any moist solids or unevaporated water droplets. As shown in Figure 44, spalling of the thick deposits begins to restrict the outlet flow area particularly where the turning vanes are closely spaced.

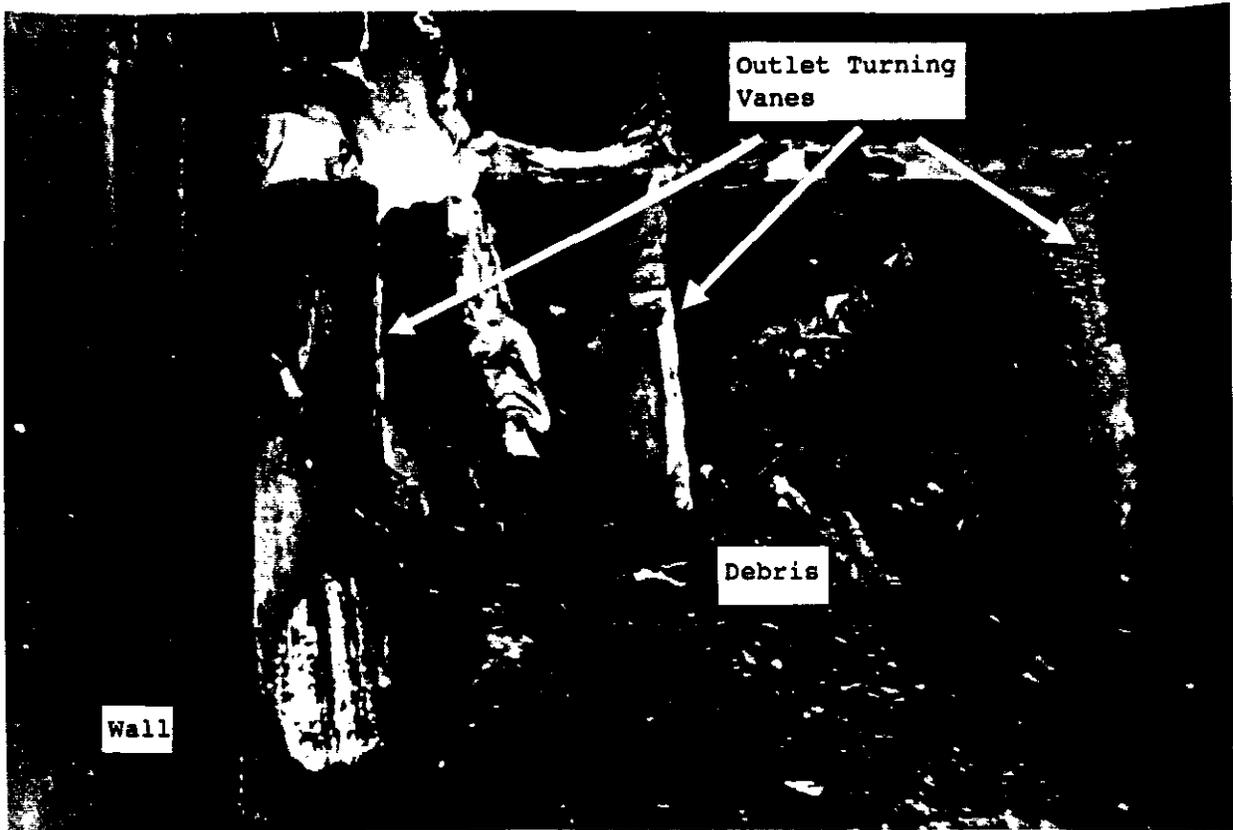


Figure 44. Outlet turning vane deposits.

Attempts were made to determine if coating the turning vanes with Teflon® sheeting would be an effective passive method of limiting the vane deposit thickness. Test sheets of Teflon® were attached to the center turning vane and to the round vane support brace as shown in Figure 45. The use of a slick surface coating increased the spalling rate and helped to limit the scale thickness as shown in Figure 46. Since the Teflon® coatings were installed near

the end of Coolside process testing, the long-term effectiveness and ruggedness to withstand fly ash abrasion and temperature cycling is not known. Vane scales will not form if there are no liquid droplets and the solids are completely dry at the humidifier outlet. .

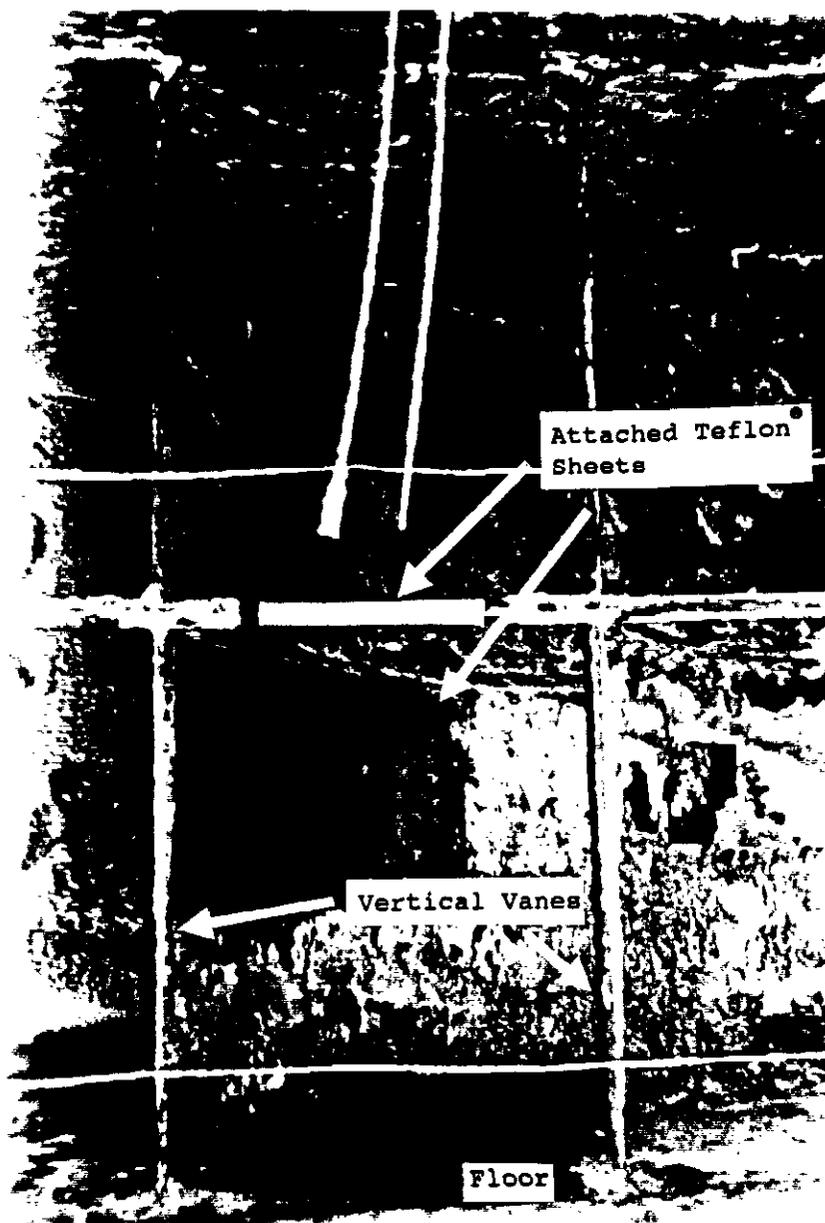


Figure 45. Outlet turning vanes with Teflon® coated surface areas.

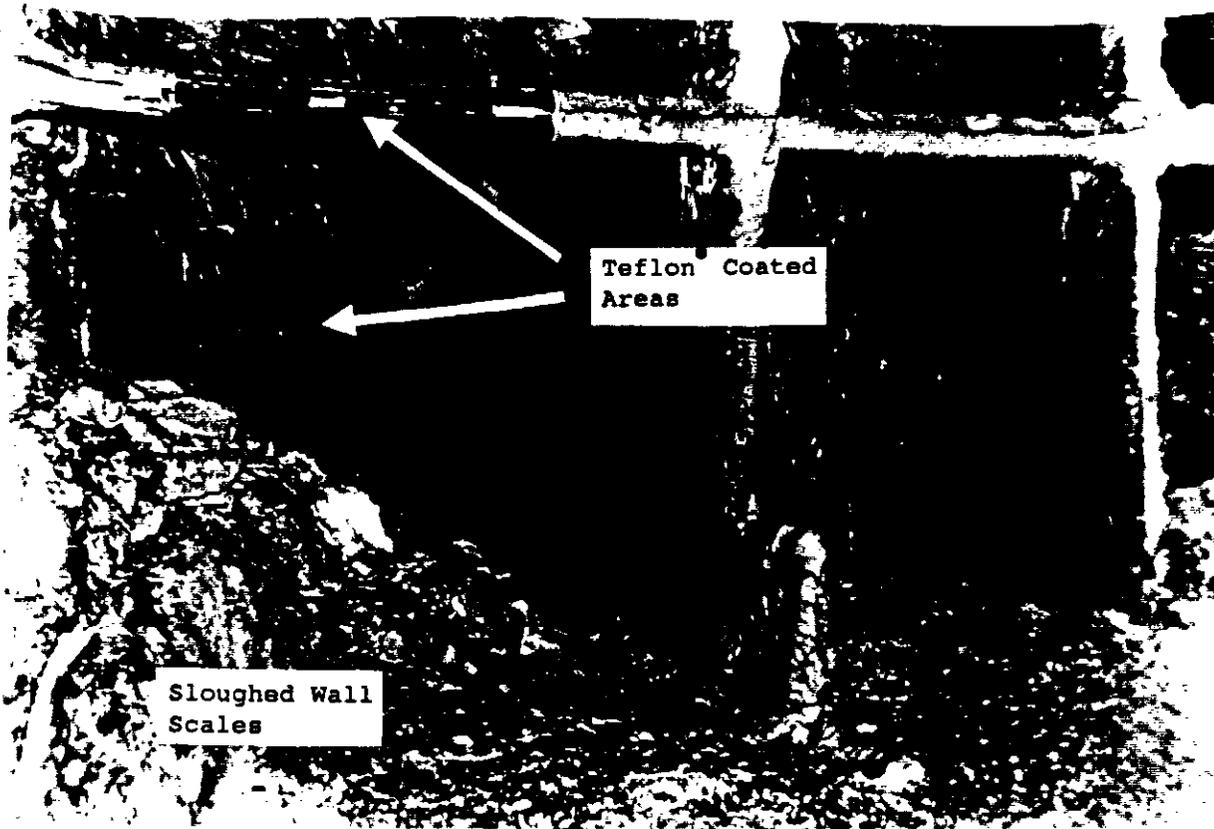


Figure 46. Outlet turning vanes showing the Teflon<sup>®</sup> coated areas following a run.

#### Thermocouple Deposits--

An array of shielded/unshielded thermocouple (TC) pairs, as shown in Figure 47, is used to measure the humidifier outlet temperatures. Using a TC array allows detection of cross duct temperature gradients which can indicate the need to adjust the water flow distribution to the atomizers. The use of shielded and unshielded TCs allows the detection of unevaporated droplets since any droplets hitting an unshielded thermocouple will cause a lower temperature reading than that obtained for the adjacent shielded TC. Complete evaporation can be assumed if the shielded and unshielded TC readings are identical and are above the saturation temperature. The construction of the original shielded/unshielded TC pairs is shown in Figure 48.

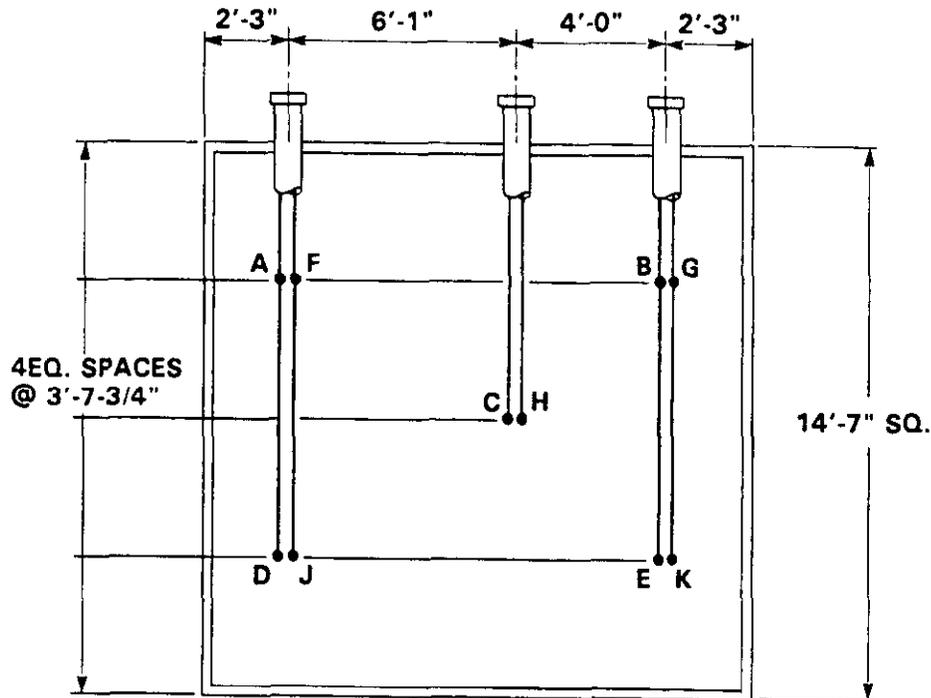


Figure 47. Location of the humidifier outlet thermocouples.

Large solids scales formed on the leading edges of the original TC probes during close temperature approach operations (see Figure 49). The scales would grow to large size and occasionally break off from the probes. This added to the floor solids accumulations at the humidifier outlet.

To minimize the size of the TC probe deposits, the TC guides were eliminated. Guy wires were used to support and position bare 1/8" diameter TCs at the humidifier outlet. Smaller TC shields made from split stainless steel tubing were attached to the sensing end of the shielded TCs. This provided a much smaller leading edge on which deposits could form. Figure 50 shows the typical condition of the modified TCs after a run. Relatively thin scales on the probe leading edges can be seen as well as on the guy wires and TC shields. Vibrations induced by humidifier gas flow turbulence combined with the thinness and flexibility of the TCs and guy wires made the modified TCs largely self cleaning.

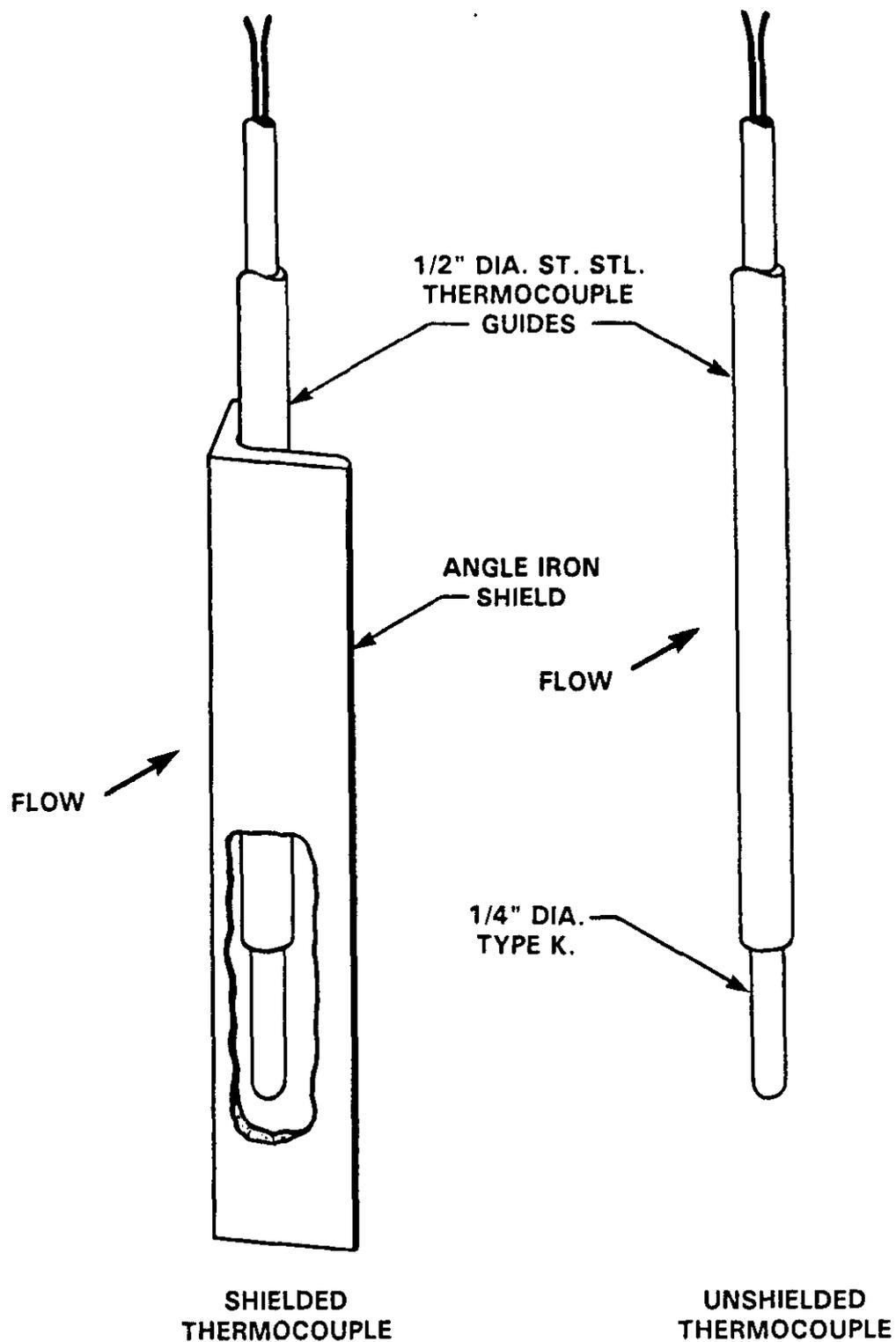


Figure 48. Construction of original shielded/unshielded thermocouple pairs.

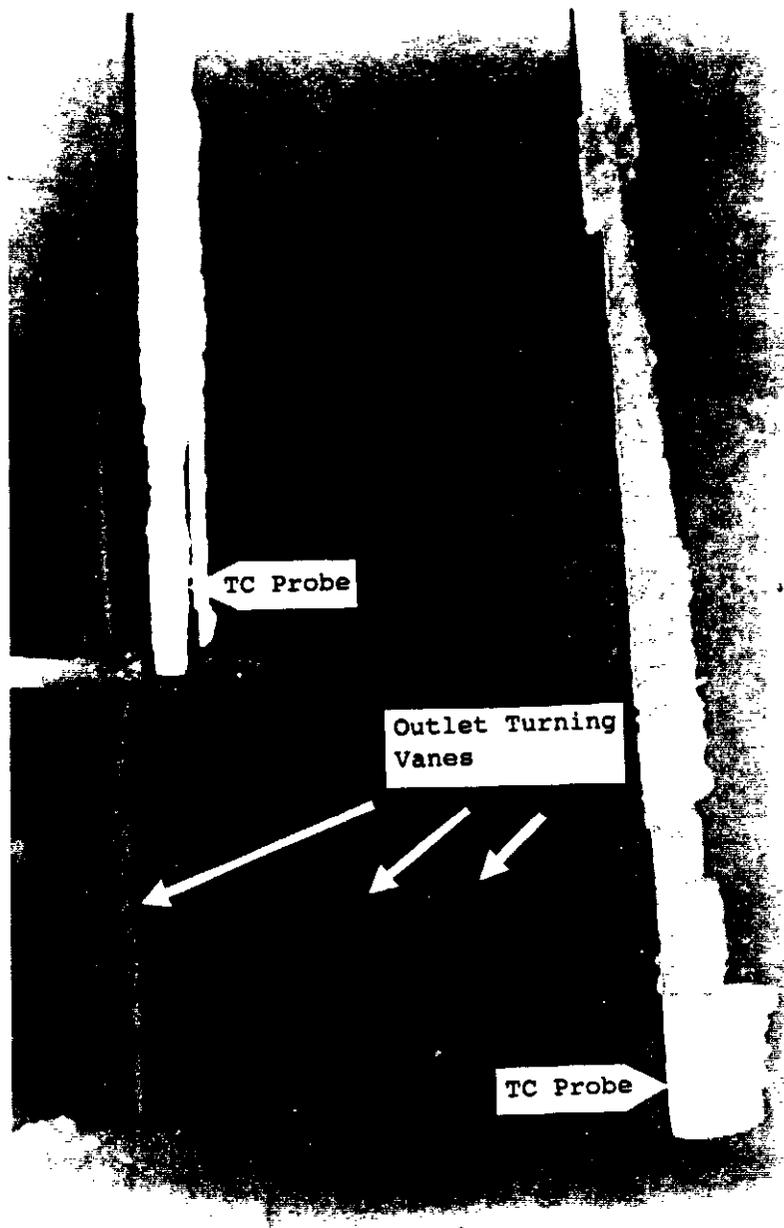


Figure 49. Humidifier outlet thermocouple deposits.

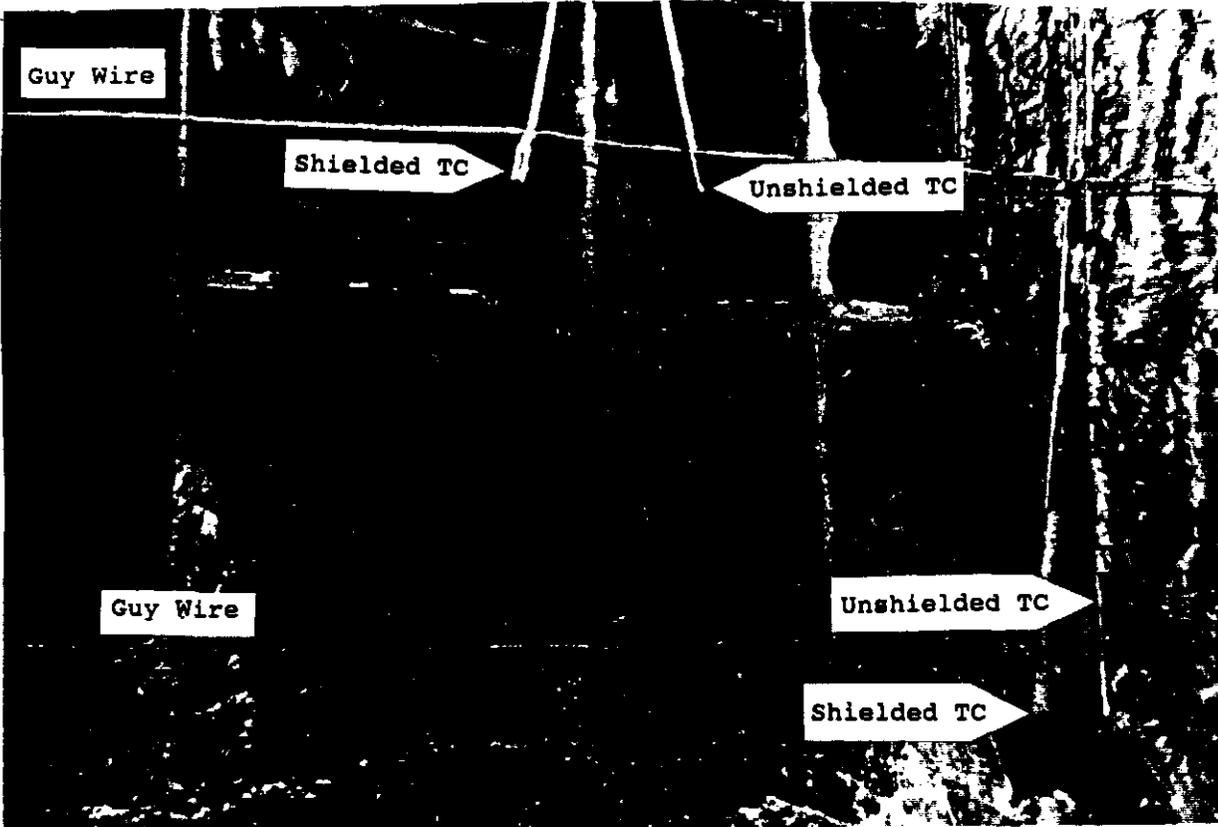


Figure 50. Modified humidifier outlet thermocouples after a run.

#### Humidifier Control

As shown in Figure 47, an array of shielded and unshielded TCs was used to measure the humidifier outlet temperatures. Originally, the water flow controller set point was reset based on the average temperature determined by the five humidifier outlet shielded TCs. This method of control worked well. However, during one Coalside test run, a process upset caused several of the outlet TCs to be buried by solids. The solids insulated the TCs from the flue gas. This resulted in an incorrect high average humidifier outlet temperature indication which caused the water flow controller to feed more water than could be evaporated in the humidifier volume. Conditions deteriorated and a large solids deposit formed at the humidifier outlet which restricted the gas flow area to about one-fourth the original area.

Several other factors contributed to the rather severe process upset. During the run, the boiler was often operated at high load, 90-100 MW. This increased the humidifier gas flow to the maximum since no flue gas was bypassed. Additionally, because of air heater leaks, the humidifier flue gas flow was about 20% higher at full boiler load than the design rate. The high flue gas rates reduced the humidifier drying time, increased the water flow to the atomizers, and decreased the atomizing air/liquid ratio. This combination of factors caused the atomizers to produce a coarser spray than design and increased the required drying time while simultaneously the actual drying time within the humidifier decreased.

To prevent similar process upset occurrences, the location of the control TCs was moved to downstream of the outlet turning vanes. At this location, gas velocity was double the humidifier velocity and the potential for solids to build up on the duct floor was much less. Additionally, the operating procedures were modified to: (1) limit the flue gas flow through the humidifier to about 850,000 lb/hr or less, (2) maintain spray fineness by keeping the atomizing air to liquid ratio at 0.5 lb/lb or greater, and (3) keep all outlet unshielded TC readings at least 10°F above the wet bulb temperature. These operating practice changes helped to insure that all the water droplets evaporated before reaching the humidifier outlet.

To implement the operating changes, a fraction of the gas flow was bypassed around the humidifier. As shown in Figure 51, the signal from the humidifier inlet duct thermal dispersion mass flow meter was used to control the position of the louvered damper in the bypass duct around the humidifier. To insure an open gas path from the boiler, the humidifier inlet isolation guillotine damper, outlet isolation guillotine damper, and the outlet louvered damper were locked open during testing. This operating control scheme permitted Coolside testing at close approach temperatures and reliable operation when boiler loads were increased beyond the design limits of the humidifier. A commercial system could

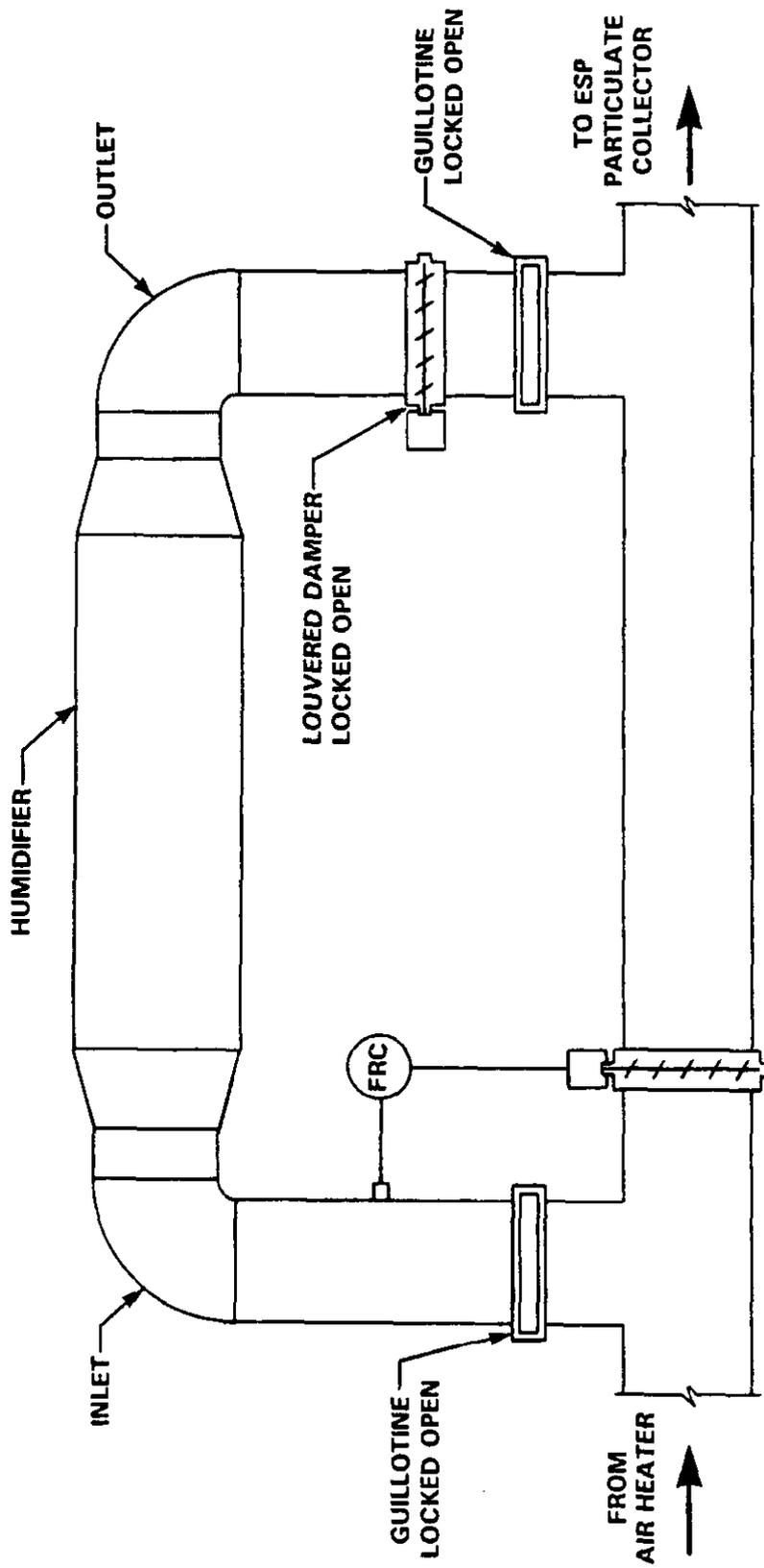


Figure 51. Flue gas flow control for Edgewater humidifier.

not likely be operated in this fashion since this limits the achievable SO<sub>2</sub> removal. This points out the need to size the humidifier for the maximum flue gas flow and temperature which may occur at the humidifier inlet (and to include adequate provision for system air leakage).

## ADDITIVE FEED SYSTEM

### General Description

Sodium hydroxide (NaOH) was chosen for the sodium additive during the Edgewater Station Coolside process tests. Sodium hydroxide was chosen for convenience and to minimize installation cost since the material could be received by tanker truck as a 50 weight percent solution which was easy to handle, store, and feed. Because of much lower price, bulk granular sodium carbonate (Na<sub>2</sub>CO<sub>3</sub>) would be preferred for commercial applications.

The final setup and control scheme for the additive feed system is shown in Figure 52. Rather than feeding the concentrated NaOH solution directly into the atomizer feed water piping, the concentrated solution was metered to the 6,000 gallon feed water storage tank. This was done to minimize the potential for precipitation of dissolved feed water solids at the mix point in the piping due to rapid pH change.

The water makeup rate to the feed water storage tank was controlled by the tank level. Water pressure to the atomizers was maintained by a pressure control kick-back flow loop between the atomizer water feed pump discharge and the feed water storage tank. A cascade control scheme was used to adjust and control the atomizing water flow rate with the water flow controller set point adjusted by the humidifier outlet temperature controller. A mixer with double blades in the feed water storage tank insures rapid and complete mixing of the NaOH with the feed water. The concentration is measured by an on-line analyzer which was in a continuously flowing water spill-back line between the atomizing water feed pump discharge and the feed water storage tank. A computer control system maintains the proper NaOH feed rate. The desired NaOH/Ca(OH)<sub>2</sub> ratio in lb/lb was manually input to the digital control system. The control system then calculated the actual feed ratio based on the humidifier water feed rate, the concentration of

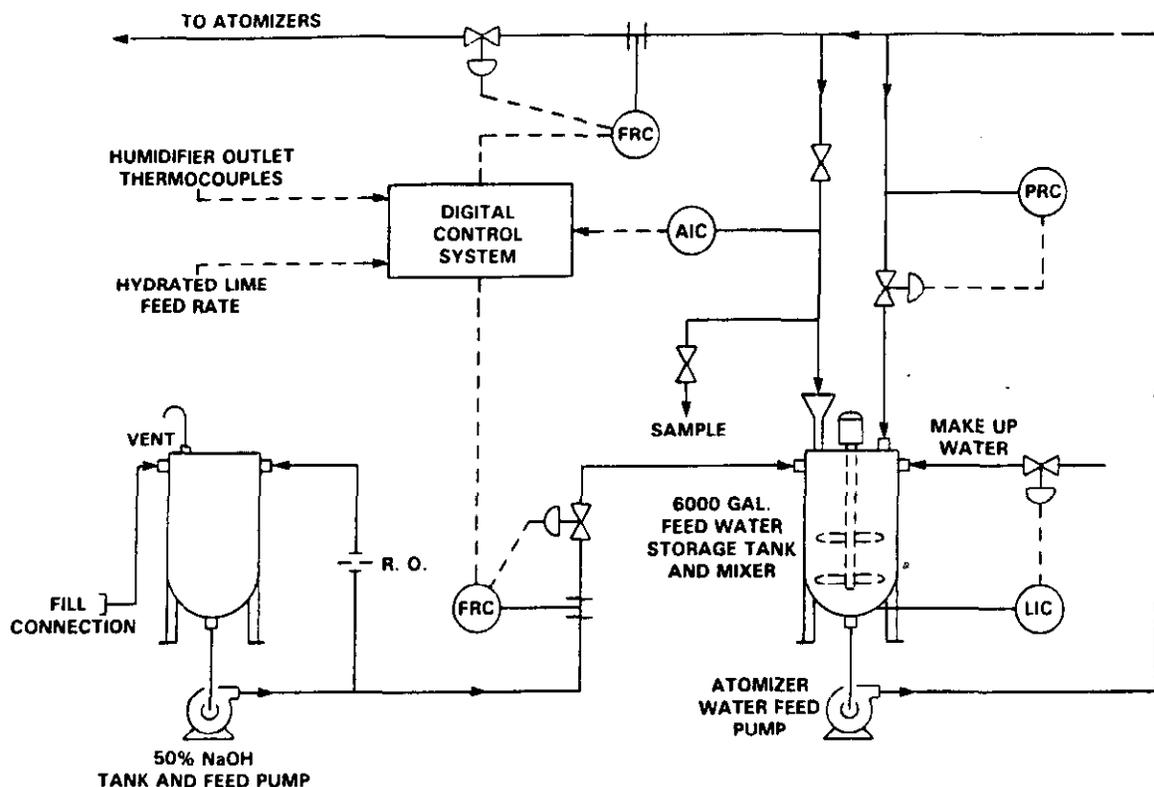


Figure 52. Sodium additive feed system.

NaOH in the feed water, and the hydrated lime feed rate. The NaOH flow controller set point is adjusted based on the calculated ratio result. Because of problems with the analyzer sodium probe, the control system was not fully operational until near the end of the Coolside test program. After the probe problem was resolved (see following section; "On-Line Ion Probe Operation"), the system worked well.

### Manual Measurement Of NaOH Concentration

To provide a check of the NaOH feed rate, Consol developed a simple method for rapid determination of the NaOH concentration in the humidification water. The procedure was to obtain a liquid sample from the humidification water header and then measure the sample conductivity using a calibrated conductivity meter. The conductivity was converted to NaOH concentration using the correlation shown in Figure 53. Submerging the conductivity probe in three separate aliquot sample solutions was sufficient to obtain a stable reading. The procedure worked well and could be completed in 5-10 minutes.

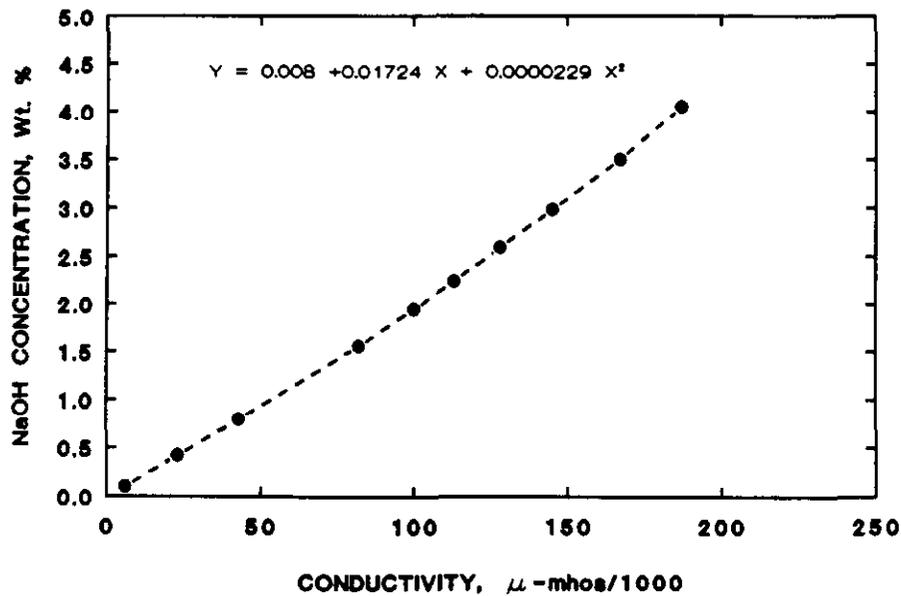


Figure 53. NaOH concentration versus conductivity correlation.

The correlation data points in Figure 53 were determined by titration of test solutions against standard acid solutions.

### On-Line Ion Probe Operation

Originally, a sodium ion specific probe was purchased to measure the sodium concentration in the humidifier feed water. The upper range of the instrument

was supposed to be capable of measuring  $\text{Na}^+$  concentrations of 100-100,000 ppm. In practice, the instrument could not be used because of extreme calibration drift.

The specific ion meter manufacturer recommended that the instrument be replaced with a newly developed conductivity meter calibrated for NaOH solutions. The new instrument was operational in February 1990. Although the instrument was in service for only ten days, it appeared to work well in the automatic control system.

### Caustic Feed Pump Operation

As shown in Figure 52, a centrifugal pump was used to maintain the NaOH header pressure. The pump kept the NaOH inventory well-mixed via recirculation through a restriction orifice line. Some pump impeller wear occurred during the testing. The cause however was not determined. Pump seal leakage was the main problem experienced. The pump seal consisted of a water-flushed lantern ring with braided Teflon® packing above and below the ring. Even after sending the pump back to the manufacturer for repair and repacking, the pump seal would leak excessively after only a few days of operation. This hampered Coolside process operations because the pump was often down for repacking. In retrospect, a seamless magnetically driven pump would have been a better choice for this service.

## RECYCLE SOLIDS FEED SYSTEM

### General Description

The general layout of the fly ash storage and recycle solids system is shown in Figure 54. A pneumatic conveying system (not shown) transported ash from ESP hoppers, boiler back pass hoppers and air heater hoppers to a 625 ton capacity ash storage silo. During weekdays, the ash was withdrawn through a dustless unloader which sprayed and mixed water and ash as the ash was loaded into dump trucks. The waste solids were then trucked to a landfill for disposal. To help maintain uniform solids flow from the ash silo, heated air was supplied through fluidizing air pads located at the bottom of the silo.

Ash could also be dumped through an alternate dry dump line into tank trucks. As shown in Figure 54, the recycle solids feed system was installed in the dry dump feed line.

The purpose of the recycle solids system was to allow recycle of hydrate containing waste solids back to the humidifier to increase sorbent utilization and thereby reduce the fresh sorbent makeup while maintaining SO<sub>2</sub> removal. The recycle solids feed rate was controlled by a variable speed rotary feeder which had a maximum speed of 17 RPM. This feeder discharged to a higher capacity constant speed (20 RPM) feeder. The constant speed feeder passed the solids to a six-inch diameter pneumatic conveying line and served as a pressure seal against the conveying air. The recycle solids were transported to a distribution bottle located just on top of the humidifier. The distribution bottle split the flow into as many as nine separate streams (normally only eight feed streams were in service for reasons discussed under previous section "Solids Injection Lance Design And Operation"). The recycle solids were fed through by two-inch diameter rubber hoses from the distribution bottle to injector pipes at the inlet of the humidifier.

A vent line between the two rotary feeders was provided to help relieve pressure across the variable speed feeder due to conveying air leakage through the constant speed feeder. Initial testing showed that pressure relief through this line was not necessary so the line was valved closed.

Conveying air was supplied by a rented, electrically driven, skid-mounted, screw-type compressor and a skid-mounted air receiver.

#### Solids Feed Rate Measurement

Because of space requirements and high costs, a gravimetric solids feed system which could provide a highly accurate measure of the recycle solids rate was not installed at Edgewater. Rather, a simple volumetric feed system similar to what would be used in commercial practice was installed.

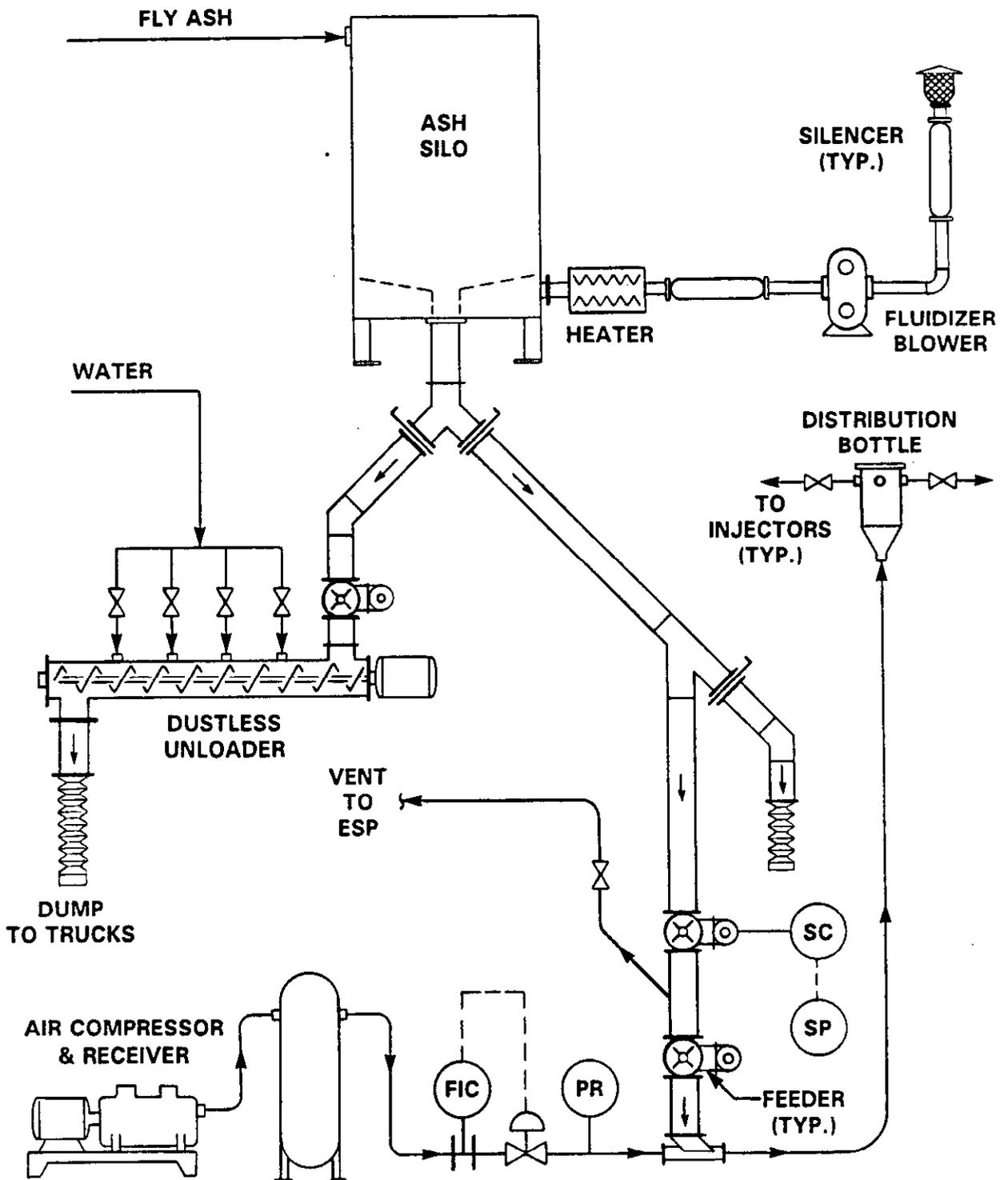


Figure 54. Recycle solids feed system design.

For this type of system, an estimate of the feed rate could be obtained from the variable speed feeder RPM and a knowledge of the feeder pocket volumes and solids bulk density. This method of estimation was subject to many errors. Factors which could affect the flow rate estimate include: incomplete filling or emptying of the rotary feeder pockets, differences in the solids density, changes in solids sneackage or flushing through the feeder due to rotor and case wear.

As part of the Coolside process tests, efforts were made to develop alternate means of estimating the recycle solids feed rate. Special short duration tests were conducted to establish the relationships between the solids feed rate and the pickup point pressure; and between the feed rate and the transfer line pressure drop (i.e., pressure drop between the pickup point and the distribution bottle).

The tests were conducted in the following fashion. First, flue gas flow was directed through the humidifier. Then recycle solids conveying air flow was established using the full capacity of the conveying air compressor (715 CFM). Recycle solids feed to the humidifier was next established and stabilized. Data on pickup point pressure and transfer line pressure drop to the distribution bottle were recorded at different RPMs for the variable speed feeder (see Appendix B, Table B-2). This allowed establishing the relationships between feeder RPM and the pickup point pressure and transfer line pressure drop (see Figure 55).

In a separate test, a special spool piece was installed in the feed line to the load cell-equipped hydrated lime day bin. At the recycle solids pickup point, a short hose was used to connect to the spool piece. Recycle solids feed was established to the day bin. When flow conditions stabilized at a given feeder RPM, the solids feed rate was determined from timed weight changes indicated by the day silo load cells (see Appendix B, Table B-3). This allowed the recycle solids feed rate to be tied to the variable speed rotary feeder RPM (see Figure 56). With this information and the data in Table B-2, correlations were developed for the recycle solids feed rate as a function of pickup point pressure and as a function of transfer line pressure drop (see Figures 57 and 58).

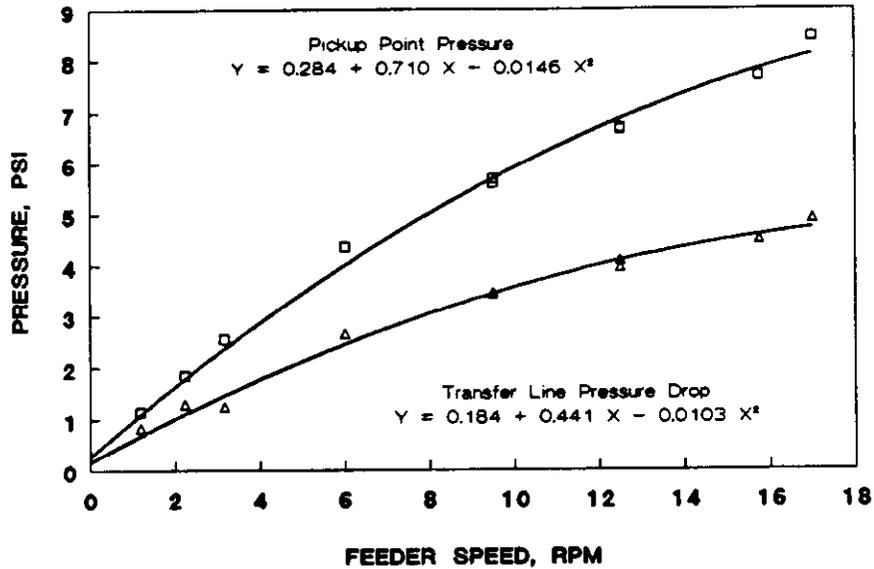


Figure 55. Recycle solids transfer line pressures (full conveying air flow condition).

Estimation of the recycle solids feed rate from the pickup point pressure or the transfer line pressure drop should be independent of the problems mentioned above: incomplete feeder pocket filling/emptying, solids bulk density changes, or feeder wear.

The use of the transfer line pressure drop appears to be the most reliable estimating method. The pickup pressure estimate proved to be unreliable during later testing when solids plugged some of the distribution bottle discharge ports. Any increase in pickup point pressure due to flow restriction translates directly into an increased calculated feed rate. The use of the transfer line pressure drop avoided this problem since the flow restrictions were downstream of differential pressure measurement.

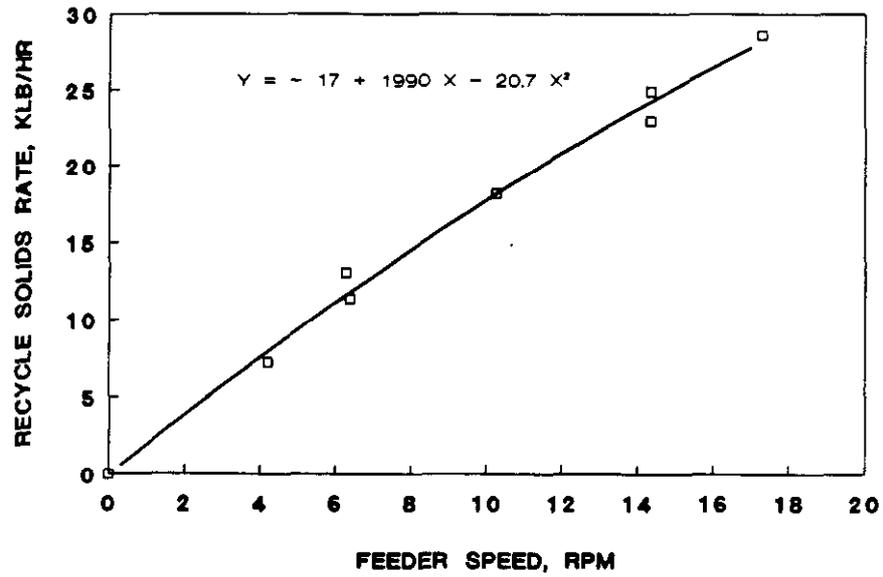


Figure 56. Recycle solids rate versus feeder speed (full conveying air flow condition).

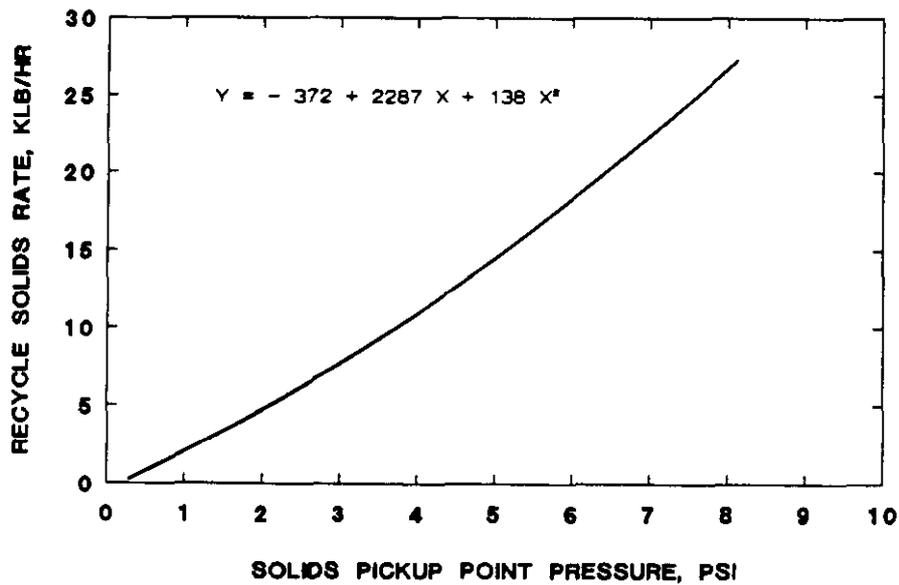


Figure 57. Recycle solids feed rate versus pickup point pressure (full conveying air flow).

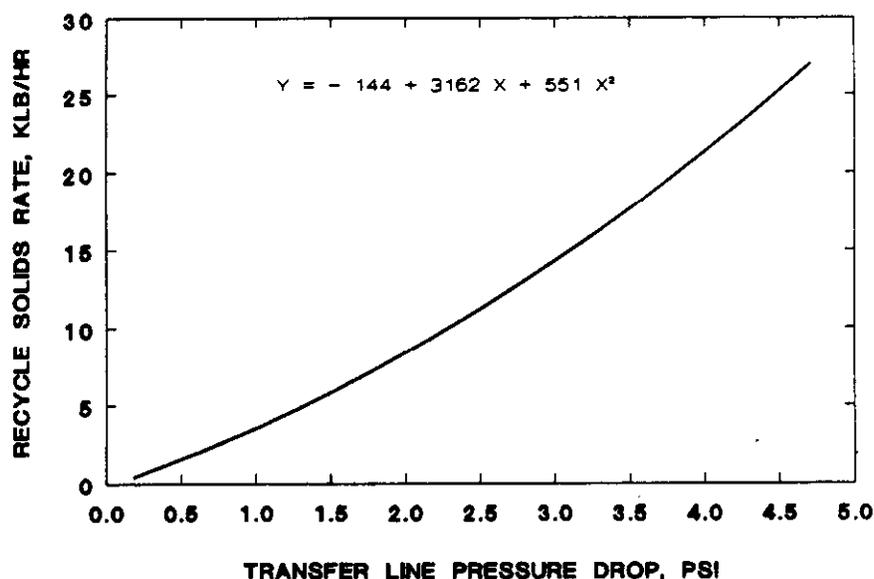


Figure 58. Recycle solids feed rate versus transfer line pressure drop (full conveying air flow).

### Transport System Problems

In general, the recycle solids feed system worked well. The operator would manually adjust the recycle solids variable speed feeder RPM set point to obtain the correct transfer line differential pressure for the desired solids feed rate. This would then be adjusted as needed.

The only problems encountered were plugging and erosion of the pipe nipples and rubber hoses at the distribution bottle outlets. The recycle solids were composed mainly of fly ash dust and fine reacted and unreacted hydrated lime. This material contained a few chunks of fused ash clinker materials which would, over time, jam and plug the 2" diameter isolation ball valves at the discharge ports of the recycle solids distribution bottle (refer to Figure 54). Sometimes turning the ball valves would break up the clinkers and clear the line. At other times, the system had to be shut down and cleaned out. Because of limited testing time and expense, no process equipment changes were made during the Coolside testing to correct permanently the problem. This problem could be

overcome by installing a screen before or after the rotary feeders to catch the coarse materials.

The recycle solids contained fly ash and were very abrasive. Severe erosion of the pipe nipples at the distribution bottle discharge ports was experienced. Holes developed in several of the 2" diameter carbon steel pipe nipples and in the rubber hoses which connect the discharge bottle to the solids injector piping. As a note of interest, no erosion problems were experienced with the hydrated lime feed system even though this system was operated over a much longer period than the recycle solids feed system. The lime tended to coat the piping and is far less abrasive than the fly ash containing recycle solids. For commercial applications, the recycle solids piping should be designed for abrasive service in the same fashion as fly ash conveying systems are designed.

## ESP OPERATION

### General Description

Particulate matter was removed from the Edgewater Station Unit 4, Boiler 13 flue gas by a Lodge-Cottrell rigid-frame ESP. Figure 59 schematically shows the system configuration and flue gas path for the Coolside process tests. After passing through the humidifier, the flue gas contained fly ash and unburned carbon from coal combustion and unreacted hydrated lime ( $\text{Ca}(\text{OH})_2$ ), calcium sulfite ( $\text{CaSO}_3$ ), calcium sulfate ( $\text{CaSO}_4$ ), and minor amounts of sodium sulfite ( $\text{Na}_2\text{SO}_3$ ) and sodium sulfate ( $\text{Na}_2\text{SO}_4$ ). During sorbent recycle tests, the solids loadings to the ESP were further increased by feeding partially reacted solids from the ash silo.

The flue gas passing through the ESP was typically at 140-145°F during operations without humidifier flue gas bypass. To prevent water condensation in the carbon steel lined stack, a flue gas reheater was provided between the ESP and the I.D. fan. The reheater used low pressure steam to provide 20-40°F of reheat. In commercial operation, the reheater would not be required if the carbon steel stack liner was coated with a protective material. Due to the limited duration of the LIMB Extension Program, this was not done at Edgewater.

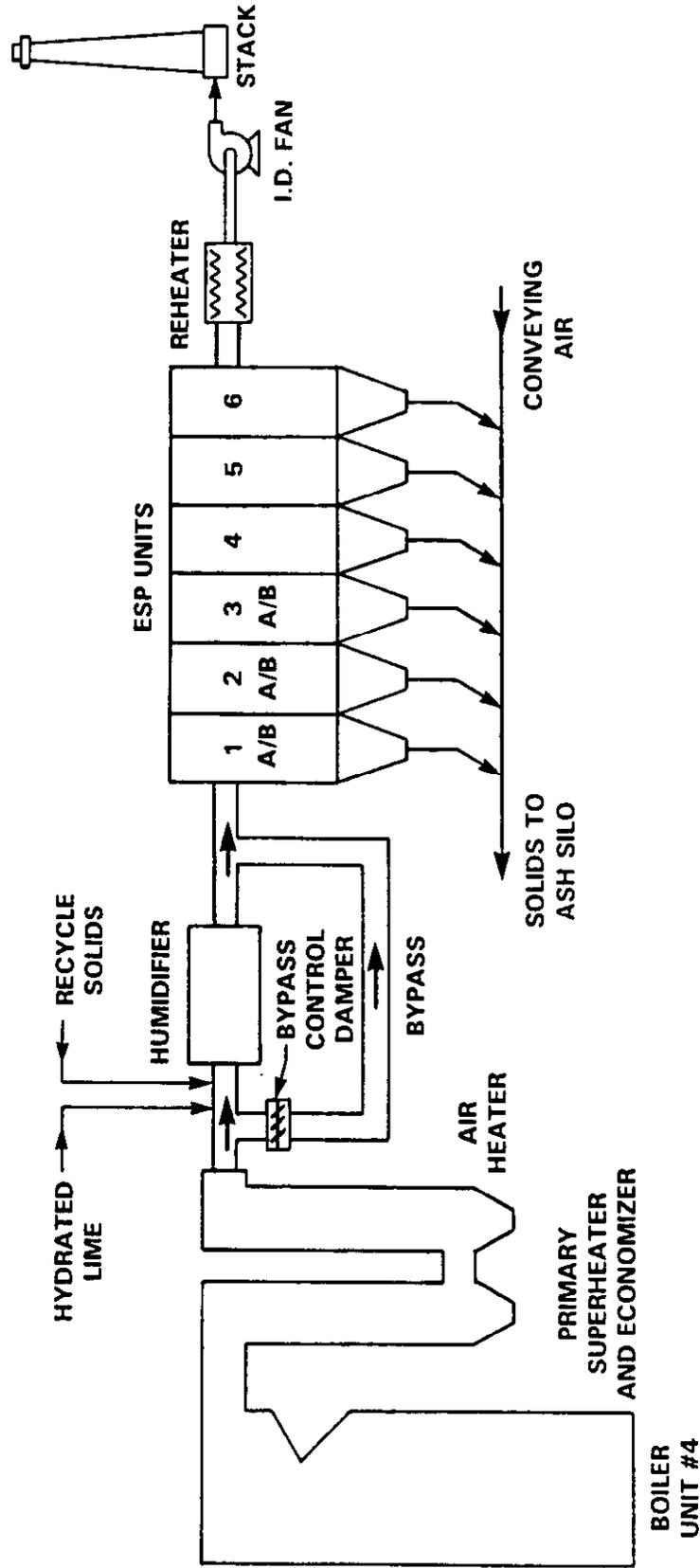


Figure 59. Edgewater Station flue gas handling system.

The Unit 4 ESP went into service in 1982, replacing an older ESP. The new ESP has six fields. The first three fields are split into A and B sections designated 1A/B, 2A/B and 3A/B; with each section having a dedicated transformer/rectifier (T/R) set. The remaining fields, with one T/R set each, are designated 4AB, 5AB and 6AB. Each field has 52,800 ft<sup>2</sup> of plate area, 44 gas passages with plate-to-wire spacings of six inches. The total plate area is 316,800 ft<sup>2</sup>. Each plate is 40 ft high and 15 ft long. Wires are eight gauge square twisted wire (0.128 in. dia.). With all fields in service, the ESP specific collecting area (SCA) is 612 ft<sup>2</sup>/1000 ACFM based on the original design flue gas flow rate of 518,000 ACFM at 280°F. The design collection efficiency is 99.38% for an inlet dust loading of 4.1 gr/ACF and for low sulfur coal (0.6 wt % S). The inclusion of a spare field and the large SCA made the Edgewater Station ideal for demonstrating and testing processes such as LIMB and Coolside which result in high ESP inlet dust loadings.

#### ESP Performance and Operating Problems

##### Wire Deposits--

During Coolside process testing, the ESP operation gradually deteriorated with time. The problem was due to the buildup of emitter wire deposits and insufficient high tension frame rapping to remove the wire deposits. The problem appeared as a continuous loss in field power and increased sparking rate beginning with the 1A/B field and slowing moving through the ESP. With time, the front ESP fields often lost all power. Cessation of Coolside operations for a few hours restored ESP operation.

Figures 60 through 68 graphically present current-voltage (I-V) data for the ESP which were obtained from Ohio Edison and other sources<sup>15,16</sup>. Where available, data for the following five operating cases are presented.

<u>Case</u>	<u>Description</u>	<u>Dust Loading</u> (Approx.) grains/SCF
1.	High-sulfur coal operation, no humidification or hydrated lime addition 9/8/87.	3-4

- |    |   |       |
|----|---|-------|
| 2. | High-sulfur coal operation with humidification and hydrated lime addition at a Ca/S mol ratio of 1.0 10/3/89.                       | 5-6   |
| 3. | Same as Case 2 but with Ca/S of 2.5 10/3/89.  | 8-9   |
| 4. | Low-sulfur coal operation with humidification, hydrated lime addition at a Ca/S mol ratio of 0.75 and recycle of ESP solids 2/8/90. | 8-12  |
| 5. | Same as Case 4 but a day later 2/9/90.  | 10-12 |

The data for the four Coolside cases (Cases 2-5) were obtained at a 20-25°F approach to the adiabatic saturation temperature. Data for Cases 2 and 3 were only taken for fields 1 A/B and 2 A/B. On February 8, 1990, and February 9, 1990, fields 1 A/B were shut down for special rapping loss tests which were conducted by ADA Technologies, Inc., and Radian for the Department of Energy (DOE).

The plots all show significant reductions in current density as a function of field strength for the Coolside operations when compared to high-sulfur coal-only operations. This current density reduction was far less for the back fields 5 and 6 than for the front fields. The curves are not typical of high resistivity problems which are often associated with high-lime content solids. High resistivity problems would be indicated by curves which eventually turn vertical or backward. Additionally, ADA's measured in situ resistivities were  $1.8 \times 10^9$  to  $4.2 \times 10^9$  ohm-cm which are not high and are in a good range for ESP operation<sup>16</sup>.

The shift in curve shape shown in the I-V plots could be due to increased plate dust layer resistivity, increases in space charge due to the presence of higher dust loadings for the Coolside process operations, increased flue gas moisture content, and in deposits on the high tension electrodes. Because of the moderate solids resistivities mentioned above, dust layer resistivity is not a likely explanation for the shift. A comparison of the data between February 8, 1990, and February 9, 1990, shows a decrease with time in the ESP energization

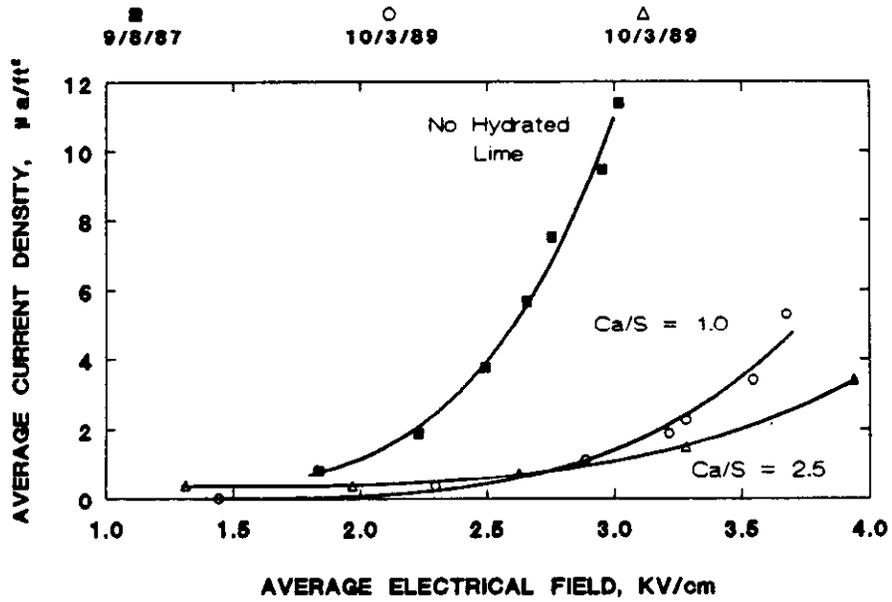


Figure 60. ESP operating characteristics -- field 1A.

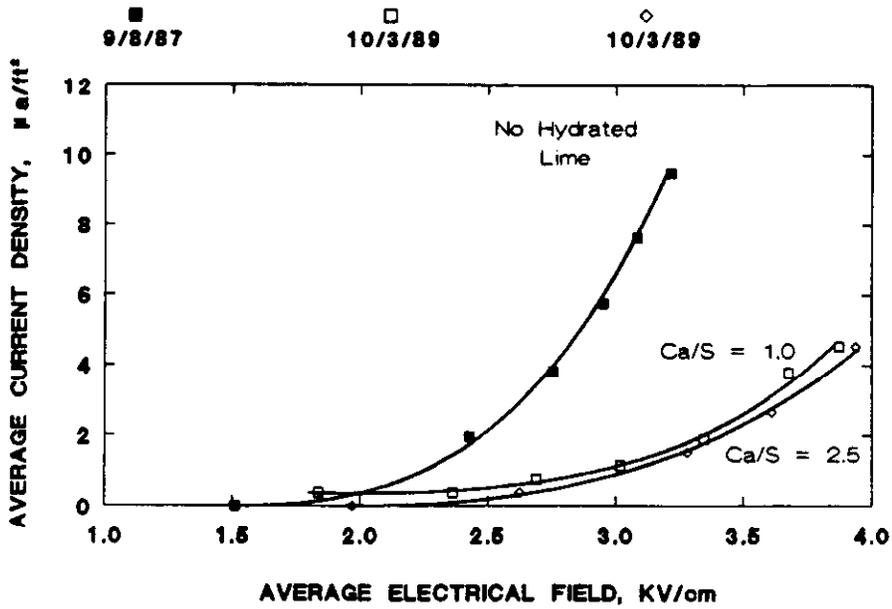


Figure 61. ESP operating characteristics -- field 1B.

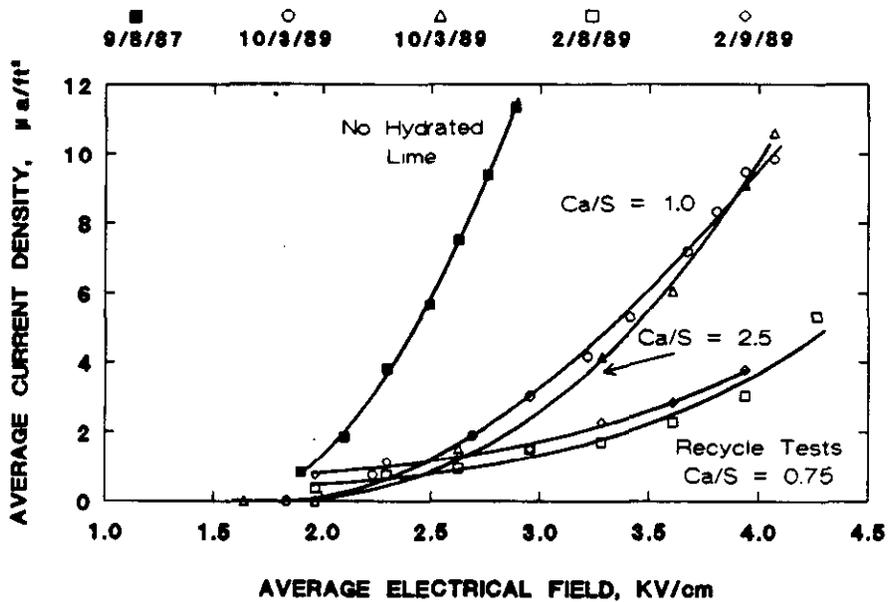


Figure 62. ESP operating characteristics -- field 2A.

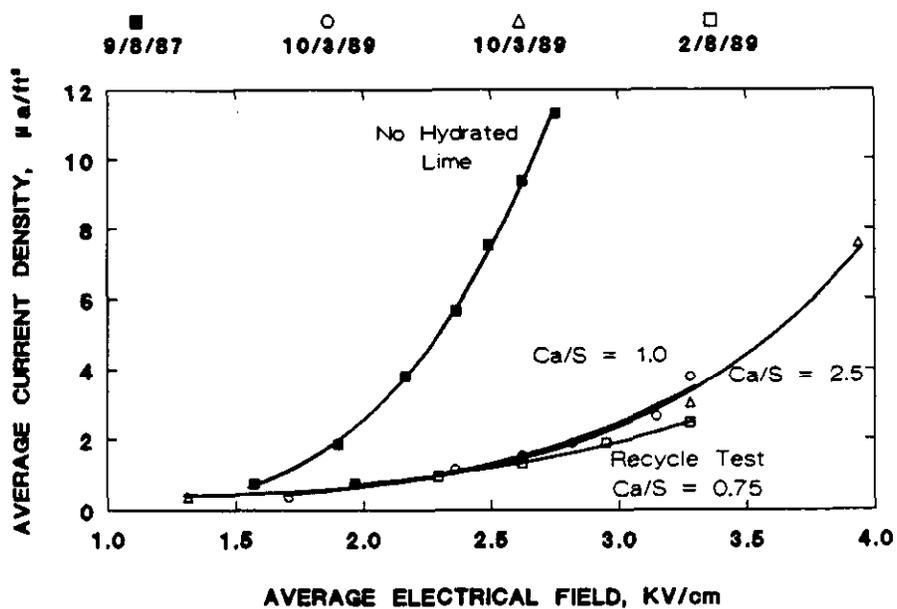


Figure 63. ESP operating characteristics -- field 2B.

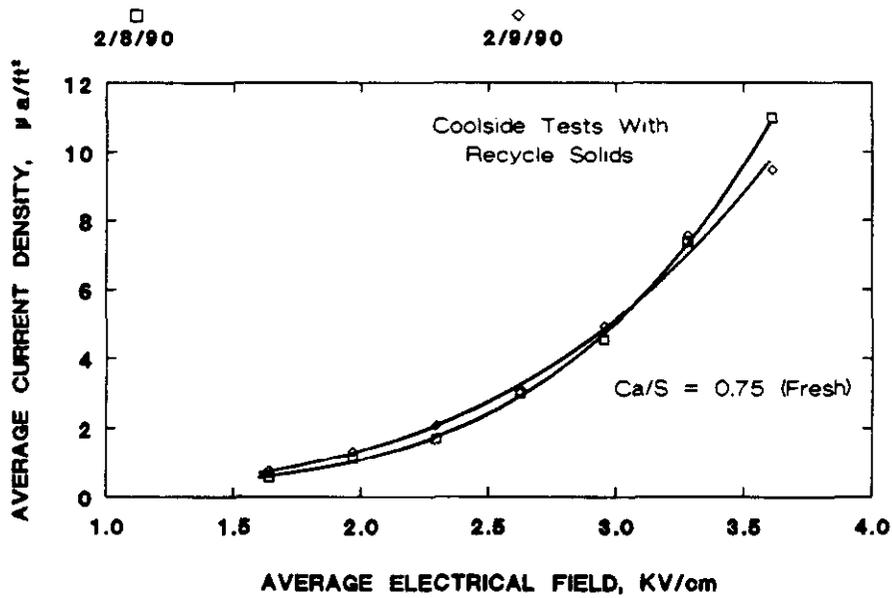


Figure 64. ESP operating characteristics -- field 3A.

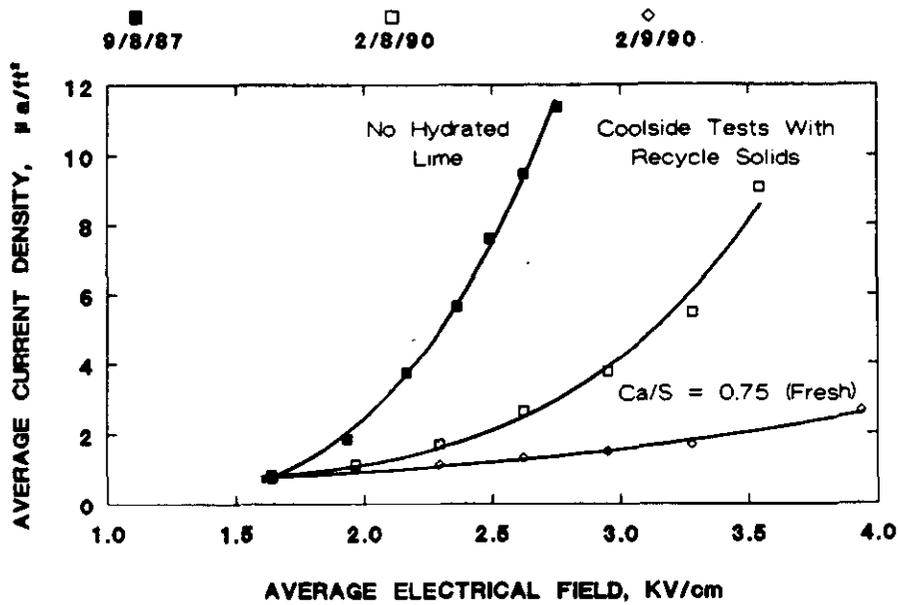


Figure 65. ESP operating characteristics -- field 3B.

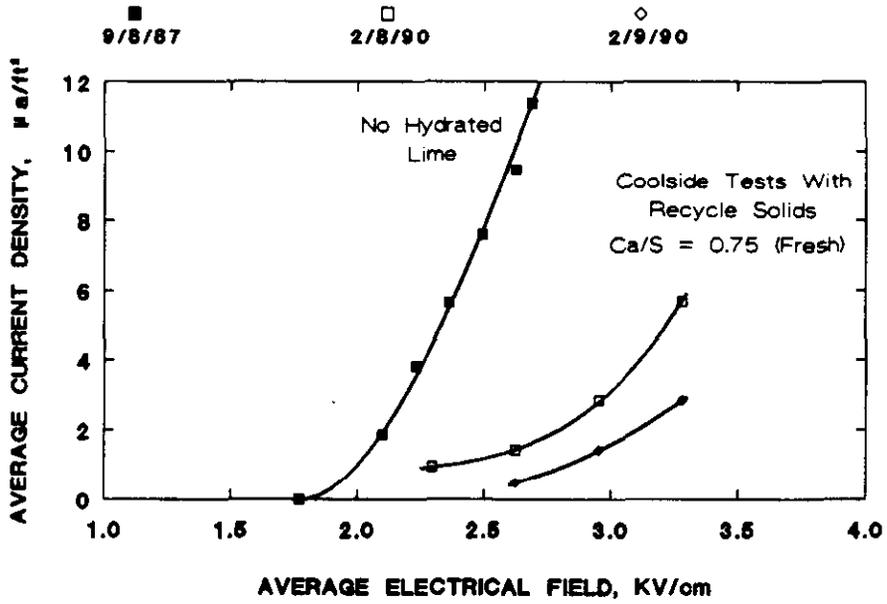


Figure 66. ESP operating characteristics -- field 4AB.

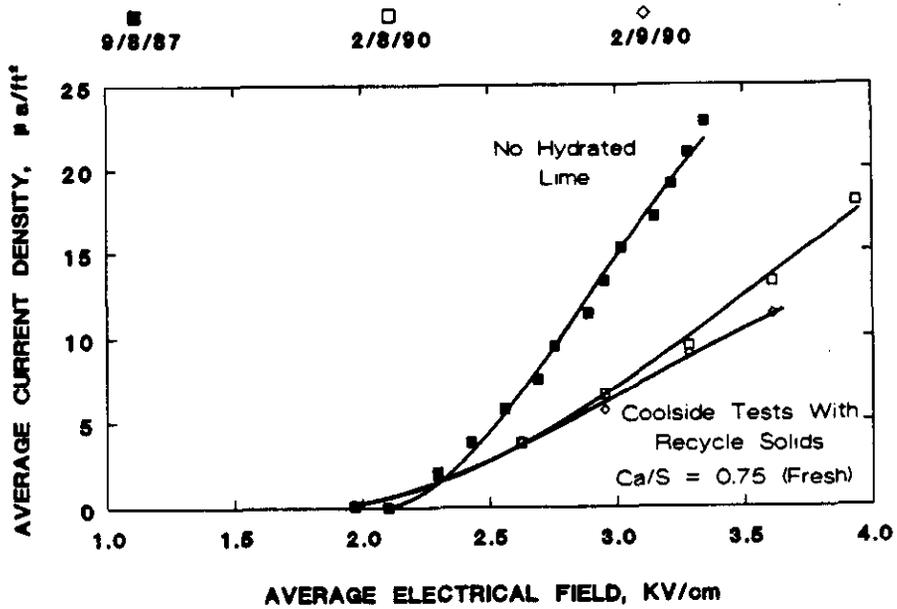


Figure 67. ESP operating characteristics -- field 5AB.

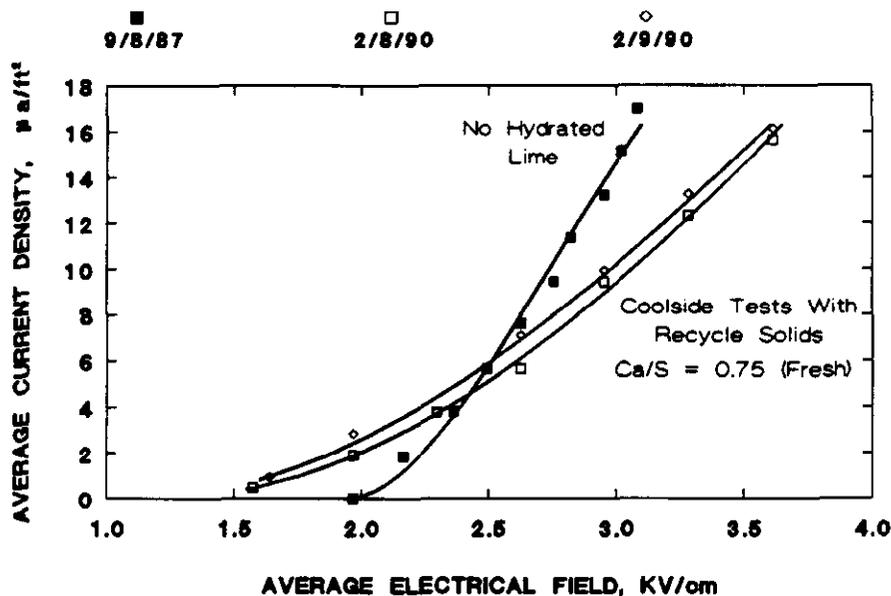


Figure 68. ESP operating characteristics -- field 6AB.

for fields 2B, 3B, and 4AB (see Figures 63, 65, and 66 respectively). Note that field 2B was completely down on February 9, 1990. The formation of wire deposits likely explains the continued decrease in the ESP energization since operating conditions were held constant over the two day period.

The ESP was inspected following the first phase of the LIMB testing in June of 1989 and following a Coolside process test in December 1989. The June inspection revealed a relatively clean ESP with only a few scattered wire deposits in the first three fields. The wire deposits (donuts) were nominally 1/2" to 3/4" in diameter. Fields 4, 5, and 6 were free of donut deposits. Collecting plates were essentially clean with typical dust layers less than 1/8" thick.

As with the June inspection, the collecting plates were found to be essentially clean during the December inspection. However, there were many more wire deposits in fields 1A/B, 2A/B and 3B. The wires which were close to the high tension frame rappers were free of donuts while the most distant wires had

donut deposits up to about 2-1/2" in diameter. Figures 69 and 70 show the typical appearance of the ESP wires close to and far from rapping points, respectively.

The presence of large wire deposits increases the effective diameter of the emitting electrodes and so would be expected to increase the corona starting voltage or field strength to an abnormally high value<sup>17</sup>. The I-V data plots of Figures 60 through 68 do not show this because the wires were not uniformly coated. Wires without deposits began corona discharge at normal voltages but current density was limited because corona discharge was reduced or non-existent from deposit covered wires. The higher sparking rate associated with the Coolside operations is explained by the decrease in wire-to-plate distance caused by the presence of wire deposits. More system weak points were formed at which spark over can occur.

Optimization of ESP operation was not a goal for the Coolside process tests. The behavior of the ESP during the Coolside tests indicates that additional evaluation is required to clearly identify causes and solutions for the ESP operating problems. Lodge-Cottrell was asked to review the available operating data and provided the following comments and recommendations.

The low temperature I-V data indicate that the T/R set voltage ratings may be limiting the power input. This can promote wire deposit formation since sparking which helps to clean wires is initially limited. A 5 kV to 10 kV increase in required operating voltage would be expected for the change in operating conditions caused by lime injection and increased water content of the flue gases. Assuming that the average available voltage from the TR sets is 55 kV, a field strength of approximately 3.6 kV/cm would be provided. Figures 60 through 68 show that up to 4 kV/cm is a typical operating level which supports the premise that the T/R set voltage is limiting.

To correct the wire deposit problem, Lodge-Cottrell recommended that existing drop rod rappers on the high tension frames be replaced with tumble hammer rappers. This would increase the rapping intensity and would be expected to restore ESP energization. Upgrading of the T/R sets may also be required

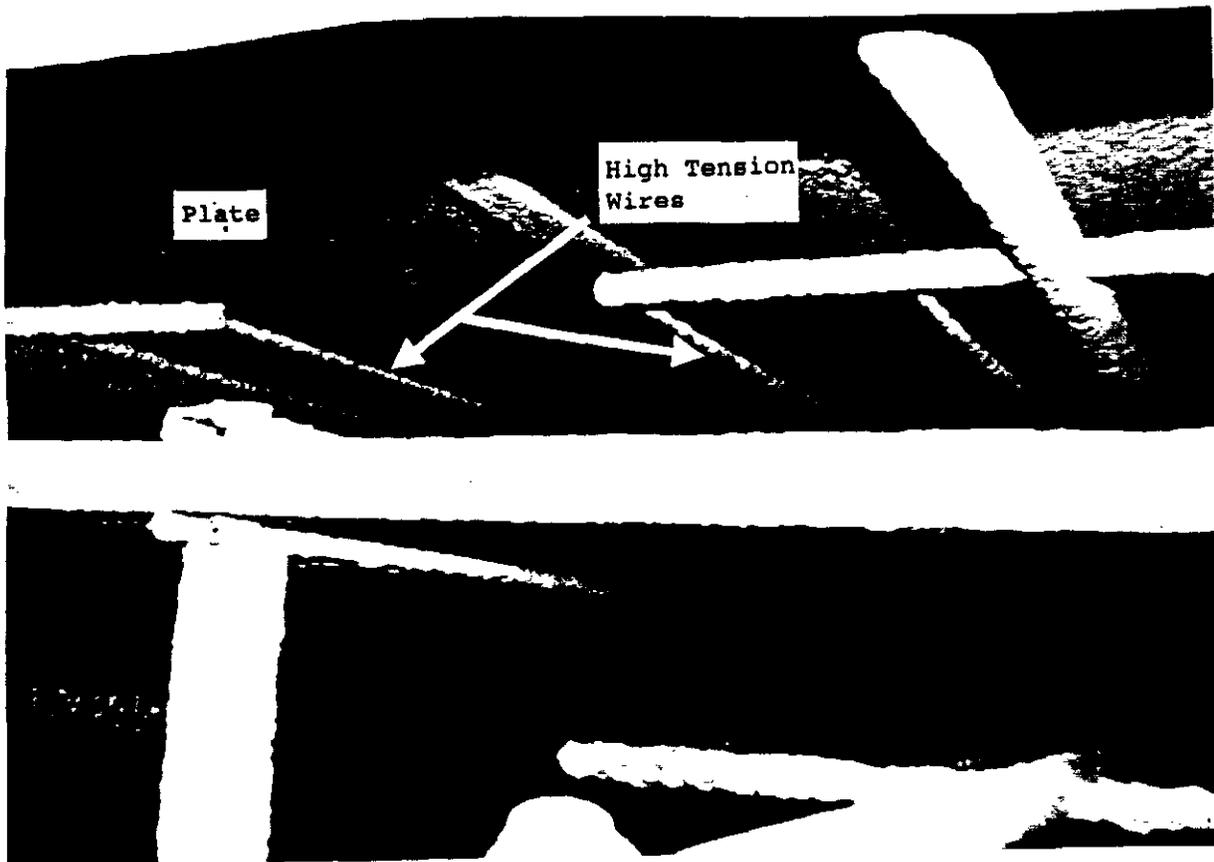


Figure 69. Clean electrodes which are near rapping points.

along with installation of high emission electrodes and high tension frame stabilizer bars to prevent frame movement. Because of expense and time constraints, these modifications were beyond the scope of the Coolside program.

#### Ash Conveying and ESP Ash Hopper Unloading--

Coolside process ash was more difficult to unload from the ESP hoppers than the normal coal fly ash. Although the Coolside waste solids were always dry, the solids tended to defluidize and pack down in the hoppers more readily than the coal fly ash. Adding a small amount of fluidizing air into the ESP hopper cones effectively improved the hopper unloading.



Figure 70. Deposits on the high tension electrodes which were furthest from rapping points.

A bottleneck was found in the ash conveying system due to the increased particulate loading associated with Coolside process operations. The design conveying rate of 36,000 lb/hr could not be maintained because of ash silo baghouse capacity limitations. The total particulate loading to the ESP was limited to about 18,000 lb/hr total solids. Installation of an additional or larger ash silo baghouse would eliminate the conveying capacity limitation. This would have allowed testing higher recycle solids feed rates to increase sorbent utilization. The equipment modification was beyond the scope of the project and was not pursued. For reasons given above, conveying system improvements would have been of little benefit without modifications to the ESP high tension rappers to improve ESP energization.

## OPERATIONS HISTORY SUMMARY

Daily operations logs of the Coolside process testing were maintained by the operating staff. These logs provide a detailed description of the equipment operations and operating problems encountered. The logs have been summarized in Appendix C.

## SECTION 5

### COMMERCIAL COOLSIDE PROCESS DESIGN AND PROCESS ECONOMICS

#### INTRODUCTION

The economics of the Coolside and the wet limestone forced oxidation (LSFO) FGD processes are compared. The evaluation is made based on capital cost and annual levelized revenue requirement for each of the two processes. The evaluation premises were developed from reviews of the Department of Energy Program Opportunity Notice (PON) (DE-PS01-88FE61530)<sup>18</sup>, and the EPRI TAG™ process economic evaluation guidelines<sup>19</sup> and from discussions between B&W and Conso1 R&D. The economic premises chosen provide a basis for comparison of the factors affecting the selection of a retrofit SO<sub>2</sub> control option.

#### PROCESS ECONOMIC EVALUATION RESULTS

The Coolside process is economically competitive with an LSFO FGD process for base load boiler operation (65% plant capacity factor) under the following baseline conditions.

1. For 1.5 percent sulfur coal, up to 350 MW<sub>e</sub> (net).
2. For 2.5 percent sulfur coal, up to 130 MW<sub>e</sub> (net).

Process sensitivity analyses show that the following factors favor the Coolside process for SO<sub>2</sub> control.

1. The Coolside process can be characterized as a low-capital cost, high-operating cost process. When compared to a high-capital cost, low-operating cost process like LSFO FGD, the economic attractiveness of the Coolside process increases with decreasing plant capacity factor.
2. The base case SO<sub>2</sub> removals are 70% and 95% for the Coolside and LSFO processes, respectively. As SO<sub>2</sub> removal requirements to achieve compliance

drop below 70%, the Coolside process becomes more economically attractive relative to wet limestone FGD.

3. A shorter remaining plant life favors the Coolside process because of the significant capital cost differential between Coolside and wet limestone FGD.

## POWER PLANT DESIGN

The economics presented in this report are based on reference plant capacities of nominally 100, 150, 250, and 500 MW<sub>e</sub> (net). The plants are located in the State of Ohio near the Ohio river. The reference plant performance assumptions are listed in Table 12. The site plan is assumed to be similar to those in DOE PON, DE-PS01-88FE61530. For the purpose of Coolside and wet FGD process layouts, all boiler sizes are assumed to be equipped with two parallel air preheaters. Each air preheater handles 50% of the flue gas flow. The flue gases exit each air preheater and flow through parallel ducts to separate ESPs.

The designs are based on eastern bituminous coals containing 1.5, 2.5, and 3.5 weight percent sulfur, as received. The fuel specifications are listed in Table 13.

**Table 12  
REFERENCE PLANT PERFORMANCE ASSUMPTIONS**

Plant Size				
MW <sub>e</sub> (net)	100	150	250	500
MW <sub>e</sub> (gross)	105	158	262	530
<b>Nominal Plant Heat Rate</b> Without FGD, Btu/kWh (net)	9830	9770	9510	9460
<b>ID Fans</b> Coolside Wet FGD	Adequate Add Booster Fan			
<b>ESP</b> Regulated Emission Rate, lb/10 <sup>6</sup> Btu Specific Collector Area, ft <sup>2</sup> /1000 acfm	0.1 400			
<b>SO<sub>2</sub> Emission Limit</b> 2.5 lb SO <sub>2</sub> /10 <sup>6</sup> Btu, 1995 1.2 lb SO <sub>2</sub> /10 <sup>6</sup> Btu, 2000	Yes Yes			
<b>Capacity Factor, percent</b> Plant Location Plant Retrofit Factors Coolside (nominal) FGD (nominal)	65 Near the Ohio River 1.3 1.3			

**Table 13**  
**DESIGN FUEL SPECIFICATIONS**

Coal Sulfur, Wt. % (AR)	1.50	2.50	3.50
Higher Heating Value (HHV), Btu/lb			
HHV (dry)	14180	14180	14180
HHV (AR)	13400	13400	13400
Coal Composition, Wt. % (dry)			
C	79.89	78.65	77.60
H	4.93	5.12	5.19
O	4.03	4.03	4.04
N	1.52	1.51	1.43
S	1.59	2.65	3.70
Ash	8.04	8.04	8.04
Total	100.00	100.00	100.00
Coal Moisture, Wt. %	5.50	5.50	5.50
Ash Content, lb/10 <sup>6</sup> Btu	5.87	5.67	5.67
SO <sub>2</sub> Potential, lb/10 <sup>6</sup> Btu	2.24	3.74	5.22

Flue gas compositions and rates depend on boiler design, fuel composition, and operating conditions. The flue gas compositions and rates used in this report are based on combustion of pulverized coal with a 140% excess air. This includes excess air to the boiler and air inleakage from the duct and air heaters. The flue gas compositions and rates are presented in Table 14. This information is included since the flue gas flow rate, moisture content, and temperature define the humidification water flow requirements for the Coolside process and the evaporation water requirements for the wet limestone FGD process. The flue gas flow rate, SO<sub>2</sub> concentration, and required Ca/S mole ratio define the hydrated lime rate for the Coolside process and the scrubber diameter, recycle pump capacity, and limestone feed rate for LSF0 FGD.

#### COOLSIDE PROCESS DESIGN

The Coolside process design is based on current FGD industry practice and the results of the Edgewater Coolside process evaluation. The Coolside process consists of four process areas: sorbent receiving/storage/preparation/feed; flue gas humidification; particulate removal and solids recycle; and waste disposal.

**Table 14  
NOMINAL FLUE GAS COMPOSITION, RATE AND TEMPERATURE**

Coal Sulfur, Wt. %	1.5	2.5	3.5
<b>Flue Gas Composition, mol %</b>			
H <sub>2</sub> O	6.16	6.35	6.43
CO <sub>2</sub> *	12.11	11.96	11.86
O <sub>2</sub>	5.89	5.86	5.85
SO <sub>2</sub> (ppm)	912	1523	2141
<b>Boiler Size, MW<sub>e</sub> (net)</b>	<b>Flue Gas Rate, MSCFM †</b>		
100	237.3	236.8	235.7
150	353.7	352.9	351.3
250	574.6	573.4	570.7
500	1154.3	1151.8	1146.4
<b>Flue Gas Temperature, °F</b>	304		
* Less than 1 percent of coal carbon lost to carbon in fly ash and flue gas CO.			
† At 80°F & 1 atmosphere.			

Sorbent/Receiving/Storage/Preparation and Feed

Two calcium sorbent options are analyzed. The first involves purchase of quicklime (CaO), which is then hydrated on-site. The second option is to purchase hydrated lime (Ca(OH)<sub>2</sub>).

Figure 71 shows the equipment required for the first option. Quicklime is normally delivered by rail to the plant and then transferred pneumatically to large-capacity storage silos. Typically, 15-30 days of bulk storage is provided depending upon site economic considerations which will be discussed in the Coolside Process Optimization section. At essentially no additional cost, truck unloading facilities at the silos are provided for emergency deliveries. From the main storage, the quicklime is pneumatically transferred to the hydrator feed bin where the lime is fed to one of two lime hydrators; one operating and one spare. The product, hydrated lime, is stored in day bins, then the lime is metered volumetrically by a powder pump and pneumatically conveyed to the flue gas humidifier.

Bulk delivery by rail is assumed for the purchased hydrated lime option. As shown in Figure 72, the hydrated lime is transferred pneumatically from the receiving site to bulk storage. From bulk storage, hydrated lime is pneumatical-

ly conveyed to a day bin and then metered volumetrically by a powder pump and pneumatically conveyed to the flue gas humidifier.

The Coolside process uses sodium compounds to increase  $\text{SO}_2$  removal and the hydrated lime utilization. In the design shown, the sodium is stored as a solution. Dry soda ash is pneumatically unloaded from trucks into a 30-day, wet soda ash storage and supply system (see Figure 73). The soda ash feed system is a packaged unit which maintains a saturated solution of sodium carbonate. The saturated soda ash solution is metered, dependent upon the hydrated lime feed rate, and the Na/Ca atom ratio setpoint, to an in-line mixer in the humidification water supply line. The soda ash supply system comes with a small dust scrubber to control dust emissions during unloading operations.

#### Flue Gas Humidification

As shown in Figure 74, boiler flue gas from both air heaters is conveyed to a single humidification chamber. In the humidifier, water containing the sodium additive is fed to an array of atomizing nozzles. High pressure air (120 psi at the atomizers) is used in dual-fluid atomizing nozzles to produce very fine water droplets (~35 micron Sauter mean diameter). Because of the spray fineness, the water evaporates completely (three seconds residence time) and quickly cools the flue gases. The rate of water addition is controlled to maintain an outlet temperature which is typically 20-25°F above the adiabatic saturation temperature.

The water atomizers selected for the Coolside process design are B&W Mark 12 nozzles or equivalent. Each nozzle is designed to operate at a 0.8-1.0 gpm throughput with an atomizing air to humidification water ratio of 0.5 lb/lb.

Hydrated lime is pneumatically conveyed to a distribution bottle or bottles where the feed stream is split into several smaller streams. The hydrated lime is then distributed into the humidifier flue gases through an array of injector pipes located in the plane of the humidification nozzles. In the humidifier, the hydrated lime reacts in the presence of high humidity with the flue gas  $\text{SO}_2$  to form  $\text{CaSO}_3$  (typically 75-85 mol percent of reacted  $\text{SO}_2$ ) and some  $\text{CaSO}_4$ .

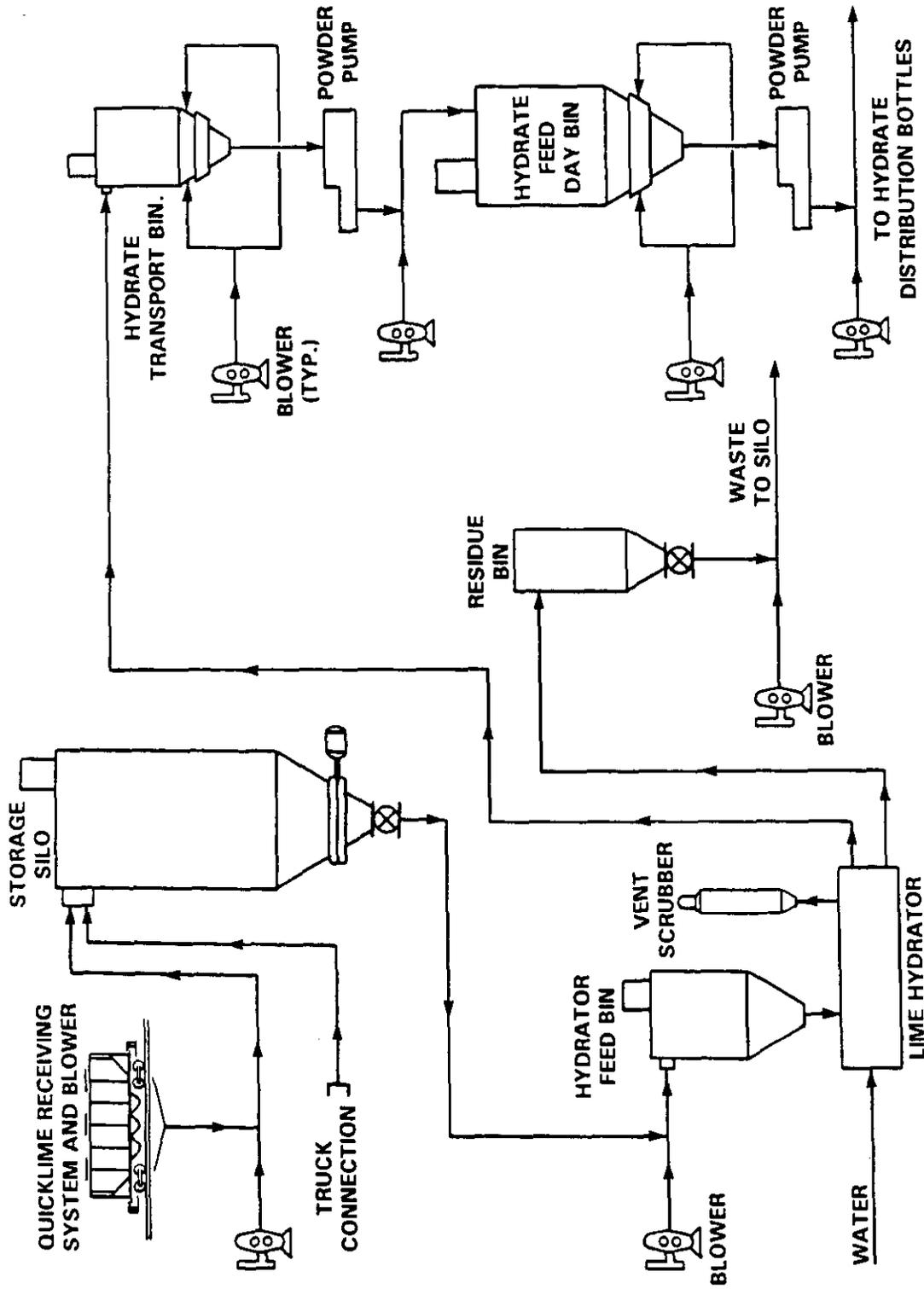


Figure 71. Sorbent feed system for on-site quicklime hydration.

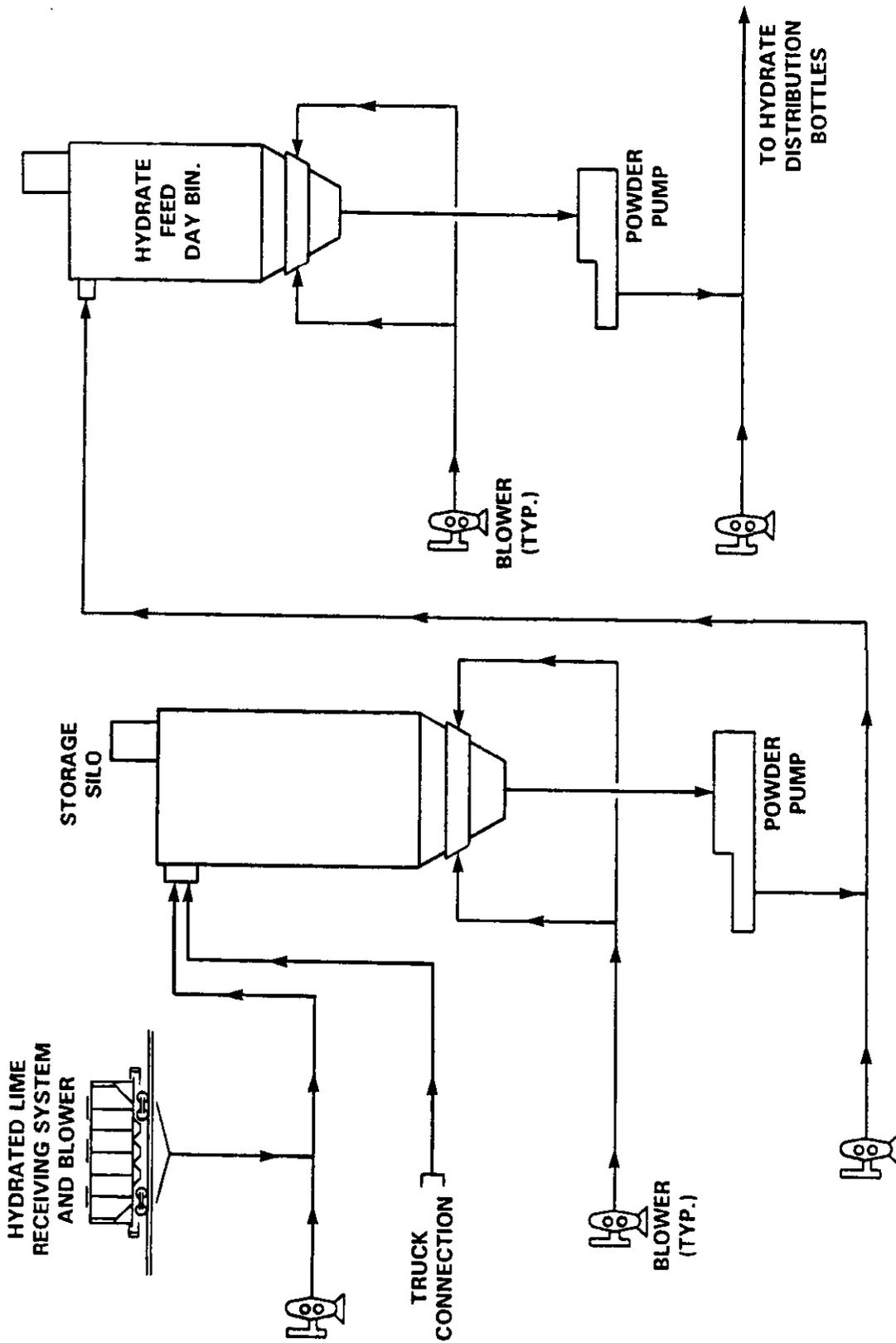


Figure 72. Sorbent feed system for purchased hydrated lime.

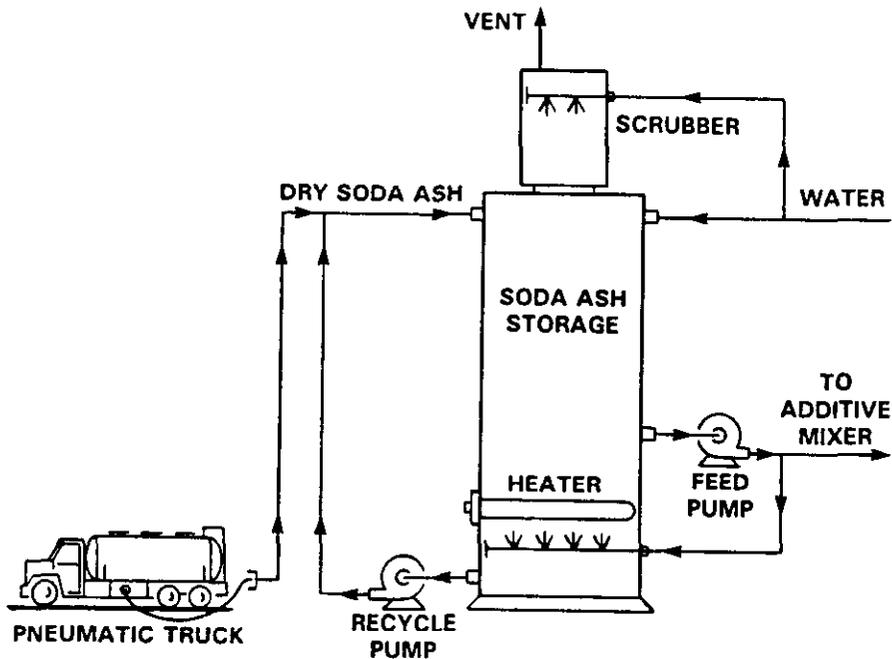


Figure 73. Coolside process sodium additive supply system.

Two key humidifier design parameters are the humidifier residence time and the inlet flue gas temperature. Based on the Edgewater demonstration, the flue gas humidifier residence time was chosen to be three seconds. To minimize the humidifier length, the humidifier cross-sectional area is set to maintain a 20 fps flue gas velocity in the humidifier. To minimize the potential for solids buildup within the humidifier, the humidifier design is vertical down flow as shown in Figure 75. A hopper is provided at the outlet of the humidifier to collect and remove wall scale and atomizer deposit debris and fly ash or sorbent which may drop out of the flue gas. The design incorporates turning vanes in all ductwork bends to minimize pressure drop and to insure a uniform gas flow profile at the humidifier inlet.

The Edgewater installation was a low velocity, horizontal humidifier design. Within the limits of existing equipment, problems with the buildup of dry flue

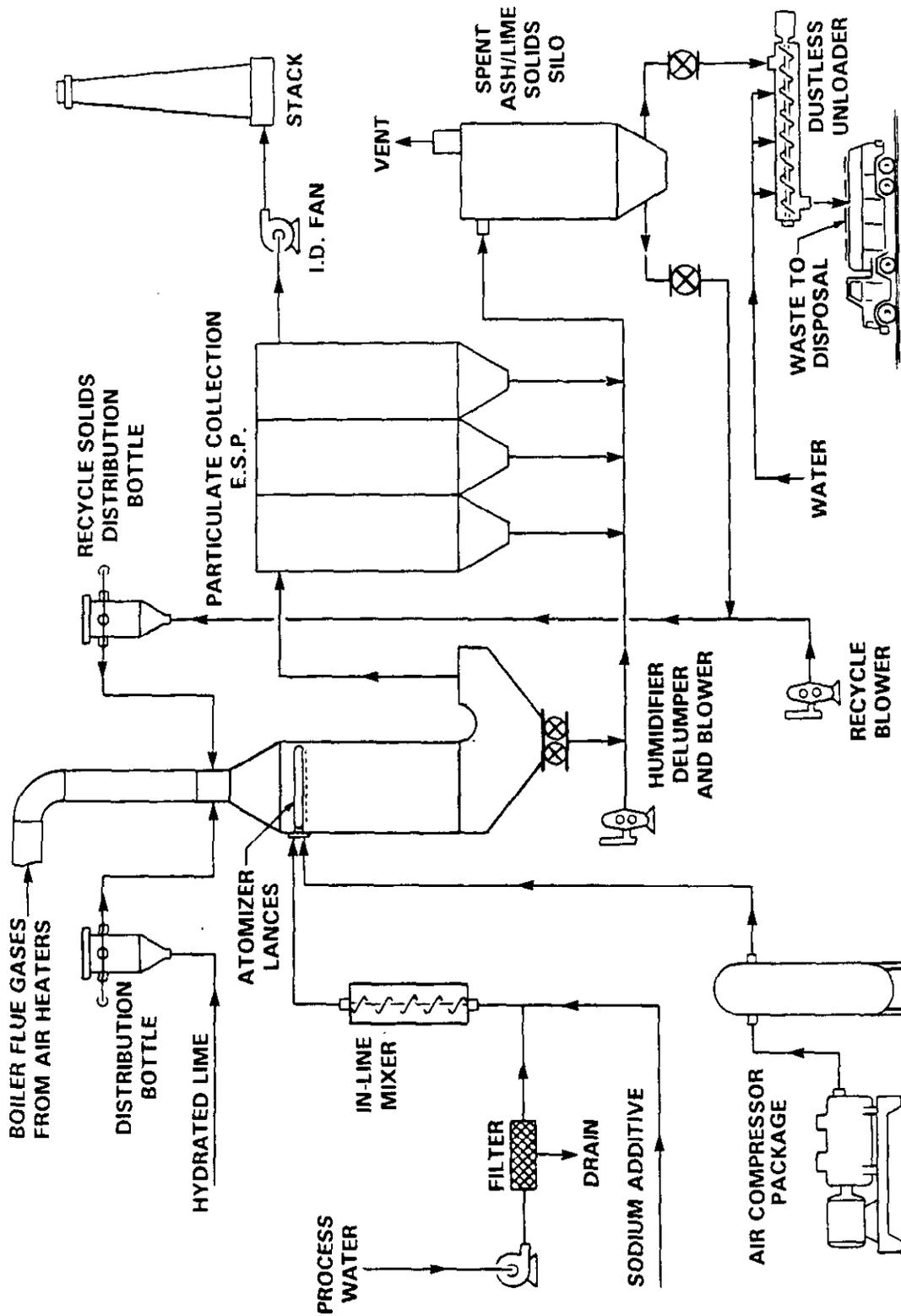


Figure 74. Coolside process humidification and recycle solids feed system.

gas solids and wall deposit fragments on the humidifier floor were not completely resolved during the demonstration program. This does not preclude site-specific applications dictating a horizontal or sloped humidifier design. Design changes such as the use of additional or more powerful floor soot blowers and the inclusion of a humidifier outlet hopper, will likely improve the operability of horizontal designs so that the humidifier can be operated long-term (i.e., six months to a year) without the need for shutdown and cleanout. In some applications, humidification in existing ducts may also be possible.

### Particulate Removal and Solids Recycle

As shown in Figure 74, flue gases from the humidifier flow to the particulate collection equipment; which in most utility applications is an electrostatic precipitator (ESP). The ESP collects and removes the fly ash, the Coolside reaction products, and the unreacted hydrated lime. The collected solids are pneumatically conveyed to a fly ash/spent lime solids silo for intermediate storage. The cleaned flue gases pass through induced draft (ID) fans to be discharged from the system through the plant stack.

To increase the sorbent utilization, recycle of the collected solids is used. Collected solids from the fly ash/lime silo are metered through a volumetric feeder into a pneumatic feed line for transfer to the humidifier. The solids are first conveyed to a distribution bottle where the flow is split and then to an array of injector pipes located in the plane of the humidification nozzles.

All reference plants used in the economic evaluation, are assumed to have: ESPs with specific collection area (SCA) of  $400 \text{ ft}^2/1000 \text{ acfm}$ , ESP particulate collection efficiencies of 99.6%, and particulate emission limits of  $0.1 \text{ lbs particulate}/10^6 \text{ Btu}$ . The SCA and expected particulate collection efficiency assumptions are based on pilot ESP test results and ESP modeling using the Consol R&D ESP computer model.<sup>20,21</sup>

For any retrofit application, the existing ESP equipment must be evaluated under Coolside operating conditions to determine if compliance with particulate matter emission limits can be maintained. Modifications to rappers or the transformer/rectifier (TR) sets may be necessary. Based on the Edgewater

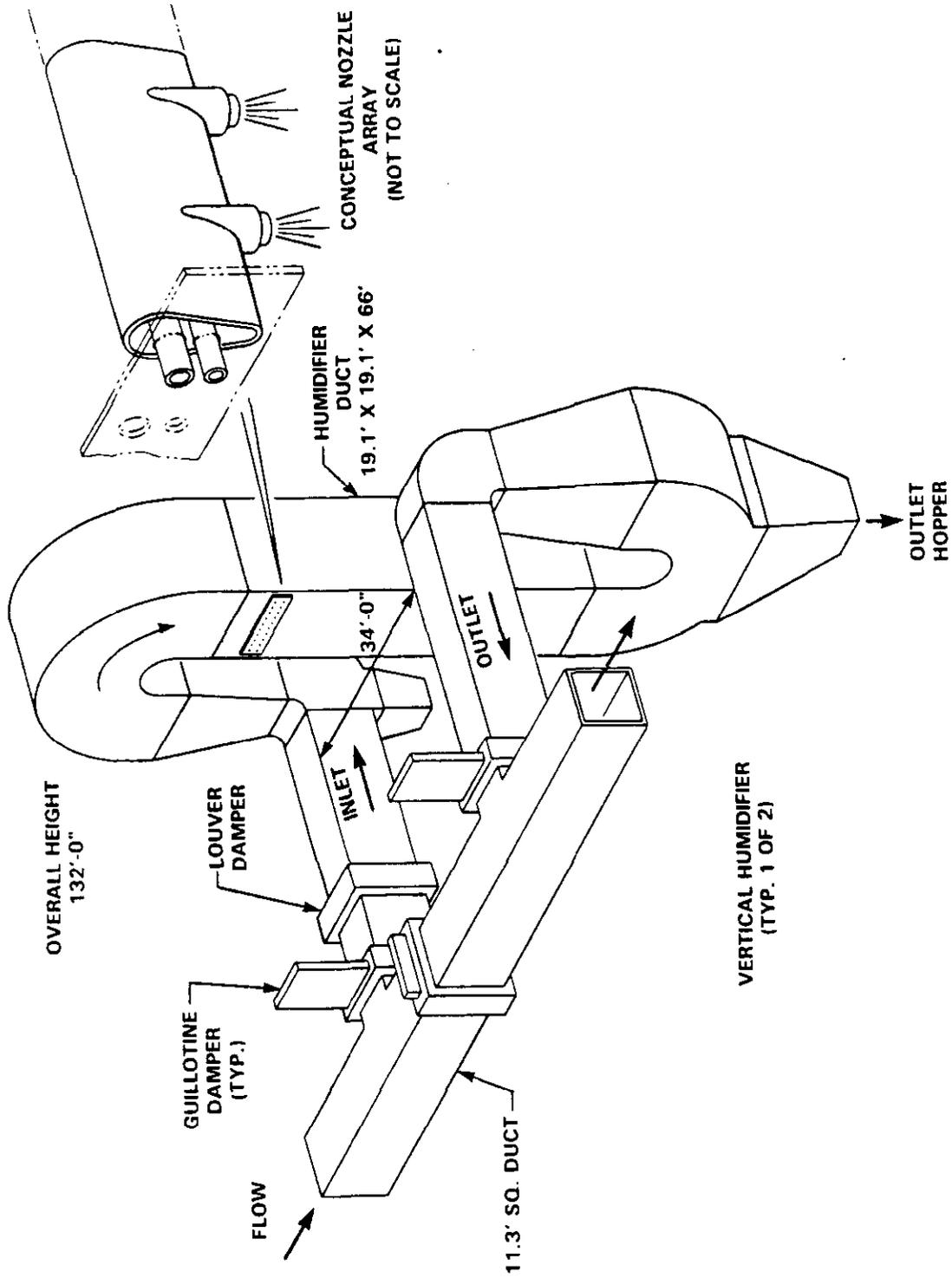


Figure 75. Vertical humidifier conceptual design -- single humidifier for 125 MW<sub>e</sub> (net) plant size.

experience, upgrading of the reference plant ESP rapping systems to increase the rapping intensity is assumed to be required for the first 300 SCA. For this study, the TR sets are assumed to be adequate.

The ESP ash hopper unloading system is assumed to be a dry design for the cost study. Costs are provided for upgrading the system to handle the increased solids flow resulting from the Coolside process based on the use of recycle. The ash storage silo capacity is assumed to be adequate and capable of storing three to four days ash and waste solids production. This was the situation at Edgewater. Modification of the ash silo unloading system is provided to allow simultaneous solids recycle and waste unloading.

### Waste Disposal

The waste material is fine and dusty. To prevent dust emissions during silo unloading, the wastes are fed through a dustless unloader where water is added to moisten the solids before discharge into dump trucks. The wastes are then trucked to a landfill for disposal.

Based on a waste management study conducted in conjunction with the Edgewater project disposal permitting, Coolside waste would be classified as a nonhazardous solid waste. For the reference plants, a clay-lined landfill with a groundwater quality monitoring system was assumed. This is consistent with recommendations in an independent study conducted by Baker-TSA, (discussed in Section 7). Depending on local/state regulations, the waste disposal site may have different construction/monitoring requirements. The capital for waste disposal landfilling is not included in the capital cost directly, but is included as a variable cost. The disposal charge per ton of waste is assumed to be adequate to cover the landfill capital, operating, monitoring, and maintenance costs.

### ECONOMIC PREMISES

Prior to the Edgewater Coolside program, Consol R&D developed computer models to estimate capital cost and total annual revenue requirements for various FGD processes including the Coolside and LSF0 FGD systems. Although these models are not deliverables under this contract, the model results were used to develop much

of the information for the economic case comparisons which follow. To check the Consol modeling methods, a comparison was made between the Consol model projected total capital requirement (TCR) for a LSFO FGD system and the costs presented in an EPRI sponsored study.<sup>22</sup> After adjusting the model inputs to the EPRI basis, the TCR costs are similar. The Consol estimate is 7% higher than the EPRI figure. This is considered excellent agreement for two independently developed equipment cost models.

The Consol models use a combination of algorithms and internal equipment data bases to estimate specific plant design equipment costs for individual equipment items or equipment packages. The process cost information was developed from internally funded proprietary FGD system design reports, vendor information on specific equipment, or FGD system quotes, and public literature sources such as the EPRI FGD system economic evaluations and commercially available equipment cost estimating manuals.<sup>24</sup> Within the model, costs are broken down into capital costs, variable costs, and fixed operating and maintenance (fixed O&M) costs. The capital costs or total capital requirement (TCR) include: total plant investment (TPI), preproduction costs, working capital, interest during construction (IDC). Variable costs include major consumables and waste disposal while fixed O&M costs include operating labor, maintenance costs, and administrative overhead.

Total annual levelized revenue requirements for an option are calculated using the levelizing techniques described in the EPRI TAG<sup>TM</sup>. Constant dollar levelization which nullifies the effects of inflation on capital carrying charges and operating costs is used for reported economics.

Installed process equipment costs (IPC) are determined by:

$$IPC = \left( \sum_{i=1}^n EC_i \times BF_i \times RF_i \right) \times CI \times SF = \$$$

*EC<sub>i</sub>* = individual equipment costs, *BF<sub>i</sub>* = bulk factor, *RF<sub>i</sub>* = retrofit factor, *CI* = process plant cost index adjustment and *SF* = site factor.

The bulk factors are assigned to each piece of equipment or equipment package to account for installation costs: labor, supervision, foundations and structure,

painting, insulation, and instrumentation. Over the years, Consol R&D has determined bulk factors based on in-house detailed equipment designs, vendor, and literature information. Depending upon the specific equipment, bulk factors typically range between 1.0-3.0 times the purchased equipment costs.

Retrofit factors are assigned to each process section to account for the difficulty of installing the new equipment in an existing facility. Typically, the retrofit factors range from 1.0 to 2.0. A value of 1.0 indicates no difference in installation difficulty between the retrofit and a new installation; while a value of 2.0 indicates a difficult retrofit with severely limited space and access. Although retrofit factors were individually assigned, the averaged retrofit factors for both the Coolside and LSFO processes ranged between 1.28-1.30. This is considered as an average or moderate retrofit difficulty with some space limitations.

Process plant cost index adjustment factors reported monthly in the Chemical Engineering magazine are used to correct the equipment costs to base year dollars (mid-1990 dollars).

The site factor adjustment corrects for regional construction costs differences. Kenosha, Wisconsin, is the base location with a 1.0 site factor assignment. For southern Ohio, the site factor was 1.06.

The total plant investment (TPI) cost is calculated by adding to the IPC the indirect field costs, home office costs, costs for bond, and all risk insurance and project contingency costs. The indirect field costs are costs for items such as field supervision, equipment rental, temporary facilities, small tools, testing and cleanup, labor burdens, crane rental, etc. These costs are 13.8% of IPC for all cases considered. Home office costs are for engineering and overhead and are 22.4% of IPC. Bond and risk insurance costs are 1.1% of IPC. No process contingency is applied to any case since the designs are assumed to be Nth plant designs. However, a project contingency of 18% of (IPC + indirect field + home office + bond and insurance costs) is assumed to cover unforeseen costs which would have been determined by a more detailed cost estimate.

Calculation methods for preproduction costs, working or inventory capital, and IDC generally follow EPRI recommendations.<sup>19</sup> Preproduction costs include 2% of TPI for equipment modifications during start-up, one month fixed O&M and one month variable operating costs. Working capital costs include 0.5% of TPI for spare parts plus one month's cost for consumables (i.e., hydrated lime, limestone, and soda ash). IDC is estimated based on EPRI guidelines including idealized construction times and uniform rates of capital expenditure.<sup>19</sup> For the Coolside process, IDC is 3.05% of TPI based on a two-year construction period; while for LSFO FGD, IDC is 6.2% of TPI based on a three-year construction.

Variable costs are costs which are dependent on process capacity and on-stream time. These costs include sorbents, additives, water, power, and waste disposal costs. The total variable costs (TVC) are calculated by:

$$TVC = \sum_{i=1}^n CR_i \times UC_i \times CF \times (8760 \text{hr/yr}) = \$/\text{yr}$$

$CR_i$  = hourly feed rate,  $UC$  = unit cost,  $CF$  = plant capacity factor

The unit costs for the economic evaluations are presented in Table 15.

**Table 15**  
**UNIT COSTS: MATERIALS, UTILITIES, DISPOSAL AND LABOR**

<u>Item</u>	<u>Unit Cost</u>
Water	\$ 0.65/1000 gallons
Lime	\$ 60.00/Ton
Hydrated Lime	\$ 60.00/Ton
Limestone	\$ 15.00/Ton
Soda Ash	\$155.00/Ton
Replacement Power	\$ 29.00/MW-hr
Flyash Disposal	\$ 7.00/Ton (dry)
Gypsum Disposal	\$ 7.00/Ton (dry)
Dry Sorbent Waste Disposal	\$ 8.60/Ton (dry)
Operating Labor	\$ 22.92/man-hr

The fixed costs for operating labor, maintenance and administrative overhead follow the EPRI TAG™ guidelines and are calculated as follows:

$$OL = ANJ \times ALR \times (8760 \text{ hr/yr}) = \$/\text{yr}$$

*OL = operating labor costs, ANJ = average number of operating positions ALR = average labor rate*

### Operating Labor Costs

For the Coolside process, the incremental operating manpower is assumed to be 2.33 men/shift if on-site hydration is used (one outside operator, one hydrator operator and one-third man control operator) and 1.33 men/shift if hydrated lime is purchased. For the LSFO FGD, the incremental operating manpower is assumed to be 2.33 men/shift (one day-shift supervisor, one day-shift lab technician, two operators on day and evening shifts, one operator on midnight shift).

### Maintenance Costs

$$AM = \sum_{i=1}^n IEC_i \times MF_i / 100 = \$/\text{yr}$$

*AM = annual maintenance costs, IEC<sub>i</sub> = installed equipment costs in area (i), MF<sub>i</sub> = maintenance cost percent in area (i)*

Depending upon service severity, values of MF<sub>i</sub> vary from 2.5% (low severity-low corrosion/erosion service) to 10% (highly abrasive or corrosive service). The maintenance costs are for both labor (40% of AM) and materials (60% of AM).

### Administrative Overhead Costs

The cost for administrative and support labor is assumed to be 30% of the operating labor and maintenance labor costs. Therefore, the administrative overhead cost (AOC) is:

$$AOC = 0.3 \times (OL + 0.4 \times AM) = \$/\text{yr}$$

The economic evaluations of process options are presented in terms of capital costs expressed as \$/kW<sub>e</sub> (net) and/or levelized revenue requirement as outlined in the EPRI TAG™. Levelized costs are expressed in terms of \$/ton of SO<sub>2</sub> removed and are presented as life cycle costs unless otherwise stated. In some cases,

the effect of short-term (10 year) levelization is considered. The economic factors needed for the financial calculations were established by B&W and are presented in Table 16. Levelized costs are in constant dollar terms which eliminates the effect of inflation on the results. Escalation of any operating cost above the inflation rate is not considered. Appendix D shows a typical economic model output for the Coolside process. Yearly consumable rates and utility requirements are presented in Appendix E.

**Table 16  
ECONOMIC FACTORS FOR COST ANALYSES**

Base Year of Estimate	1990 Mid Year	
Book Life, years	20	
Tax life, years	15	
Discount Rate	6.1	
Constant Dollar Levelizing Factors	<u>Life Cycle</u>	<u>Short-Term</u>
		<u>10 yr</u>
Expenses	1.000	1.000
Capital	0.118	0.134
Construction Period, years		
Coolside process	2	
LSFO FGD	3	

#### COOLSIDE PROCESS OPTIMIZATION

The Coolside process has the potential of being a low-cost retrofit FGD option for up to 70% SO<sub>2</sub> removal. Factors which affect the overall system capital or operating costs are discussed in this section. The intent is to show how system costs can be minimized through the choice of equipment.

#### Sorbent Receiving/Storage/Preparation and Feeding

##### On-Site Hydration Versus Purchase of Hydrate--

As shown in Table 15, the delivered costs of hydrated lime and quicklime are identical on a weight basis. This is a typical situation based on vendor quotes. However, the question remains; based on moles of Ca(OH)<sub>2</sub> delivered to the process, which is cheaper, hydrate quicklime on-site or purchase hydrated lime? The capital costs for an optimized Coolside process with on-site hydration versus one with hydrate purchase are shown in Figures 76, 77, and 78 for coals with

sulfur levels equivalent to 2.2, 3.7, and 5.2 lbs SO<sub>2</sub>/10<sup>6</sup> Btu, respectively. Clearly, adding on-site hydration equipment significantly increases the overall process capital costs for all cases. The capital cost for on-site hydration is \$30-\$35/kW for the smaller scale (100 MW) plants and \$5-\$8/kW for the larger scale (500 MW) plants. Because of reduced equipment requirements, purchase of hydrated lime is always the least capital cost alternative.

Levelized cost comparisons reported in terms of \$/ton of SO<sub>2</sub> removed are shown in Figures 79 to 81. The economic assumptions are \$60/ton for sorbent (quicklime or hydrate) cost and a 65% capacity factor. For the 2.2 lb SO<sub>2</sub>/MM Btu coal case, the purchase of hydrate is cheaper for all plant sizes. Crossover points where on-site hydration is favored are at plant sizes of about 250 MW<sub>e</sub> and 190 MW<sub>e</sub> for the medium- and high-sulfur coal cases, respectively. However, where savings occur for on-site hydration, the savings are always small compared to the levelized costs for hydrate purchase. Because the economics do not show a strong justification for on-site hydration at current sorbent pricing levels, purchase of hydrated lime was selected as the basis for the Coolside versus LSFO FGD economic comparisons.

The impact of changing the hydrated lime delivered cost on the levelized costs is presented in Figure 82. A delivered sorbent price change of ±\$10/ton, results in a levelized cost change of \$25 and \$28/ton of SO<sub>2</sub> removed for the low- and medium-sulfur coal cases, respectively.

#### Sorbent Storage Costs--

Normally, FGD systems would be designed to have 30 days sorbent storage. Less sorbent storage may be acceptable in situations where sorbent supply is very reliable, the supplier is near the sorbent source, or in retrofit applications where space or plant life may be a limiting constraint. The effect of on-site storage capacity on capital cost was evaluated. The capital costs savings for reducing hydrated lime sorbent storage from 30 days to 15 days or to 7 days are presented in Figures 83 and 84, respectively. For the 15 day on-site sorbent

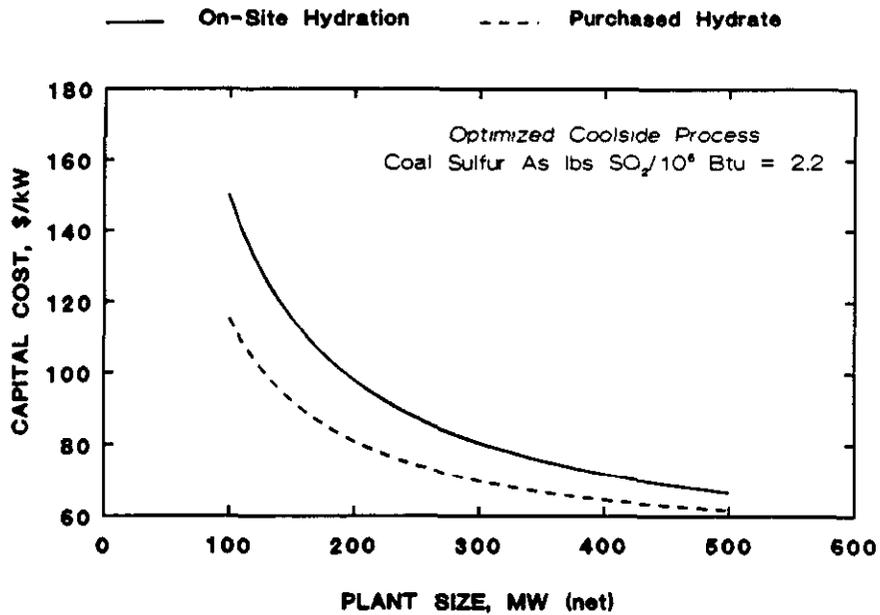


Figure 76. Coalside process capital cost comparison for on-site hydration versus purchase of hydrated lime -- (low-sulfur coal case).

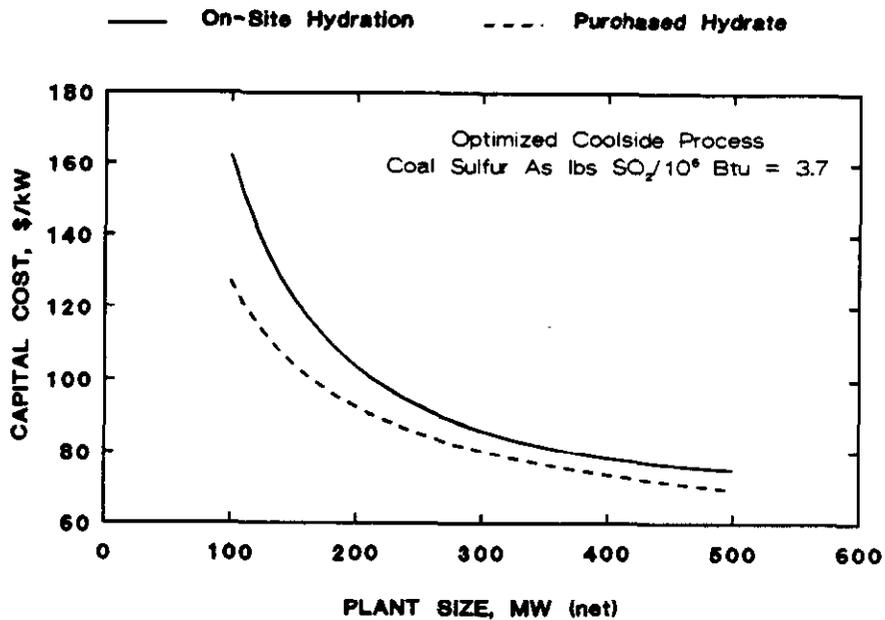


Figure 77. Coalside process capital cost comparison for on-site hydration versus purchase of hydrated lime -- (medium-sulfur coal case).

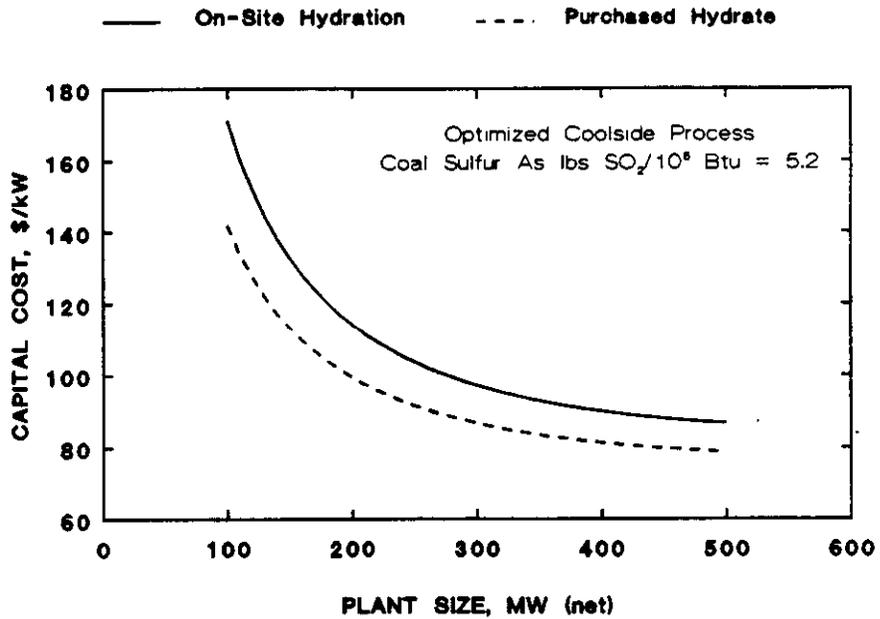


Figure 78. Coalside process capital cost comparison for on-site hydration versus purchase of hydrated lime -- (high-sulfur coal case).

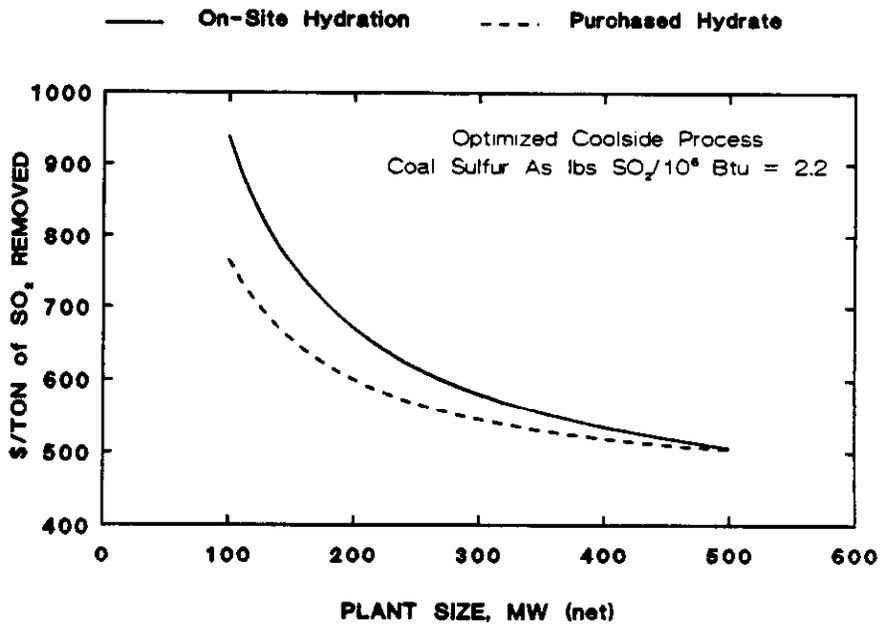


Figure 79. Coalside process constant dollar leveled cost comparison for on-site hydration versus purchase of hydrated lime -- (low-sulfur coal, \$60/ton quicklime or hydrate, 65 percent boiler capacity factor).

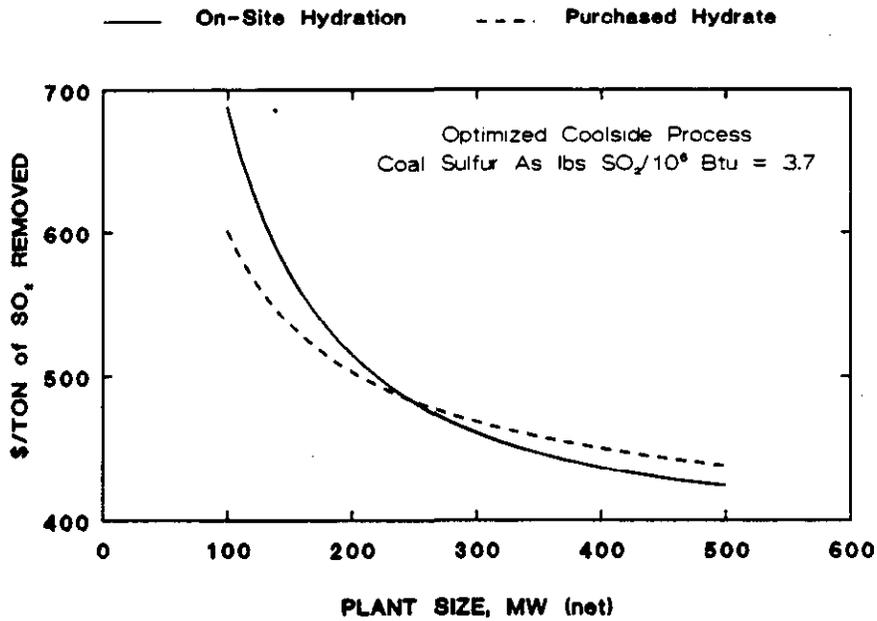


Figure 80. Coolside process constant dollar levelized cost comparison for on-site hydration versus purchase of hydrated lime -- (medium-sulfur coal, \$60/ton quicklime or hydrate, 65 percent boiler capacity factor).

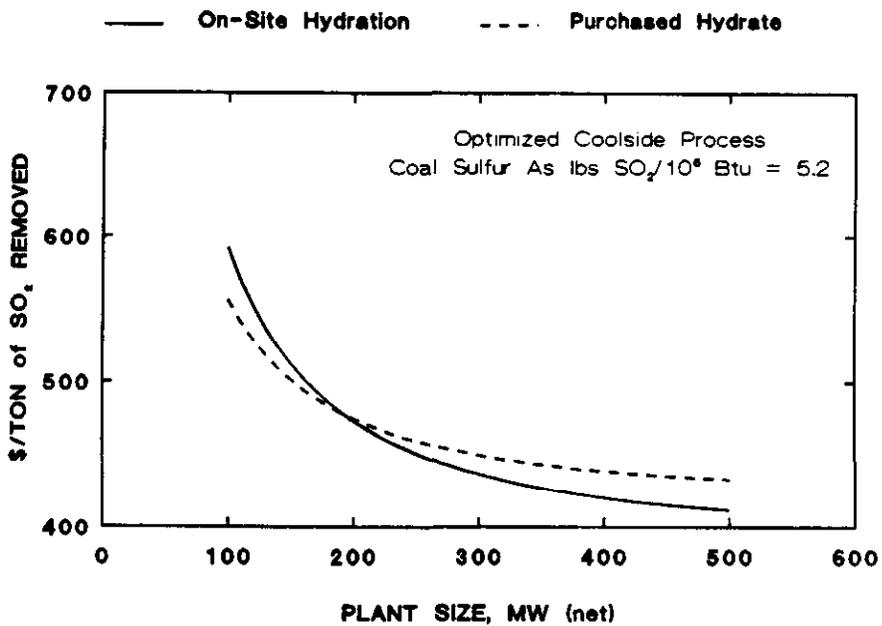


Figure 81. Coolside process constant dollar levelized cost comparison for on-site hydration versus purchase of hydrated lime -- (high-sulfur coal, \$60/ton quicklime or hydrate, 65 percent boiler capacity factor).

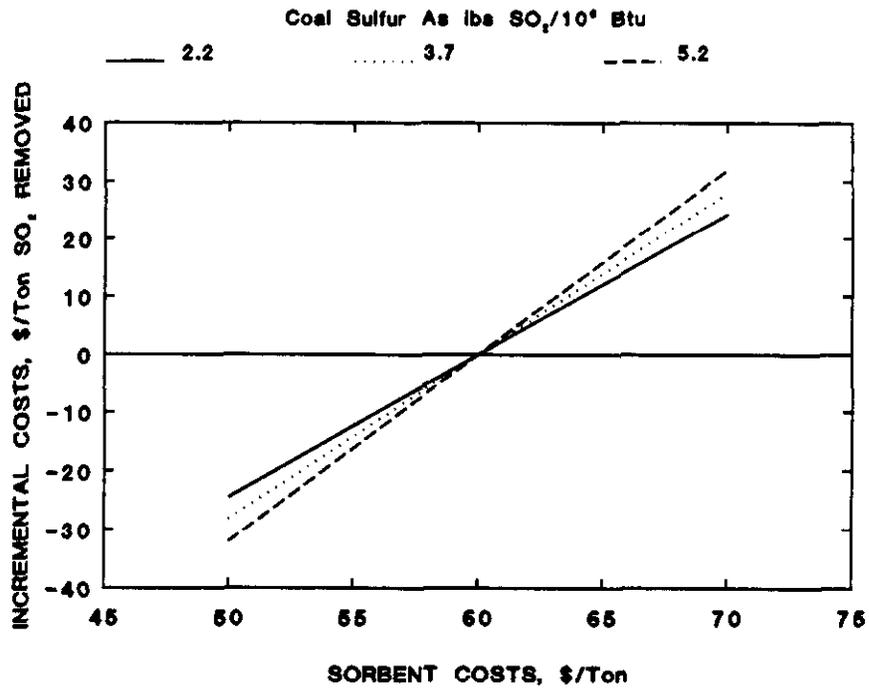


Figure 82. Effect of hydrated lime purchase price on Coolside process levelized costs.

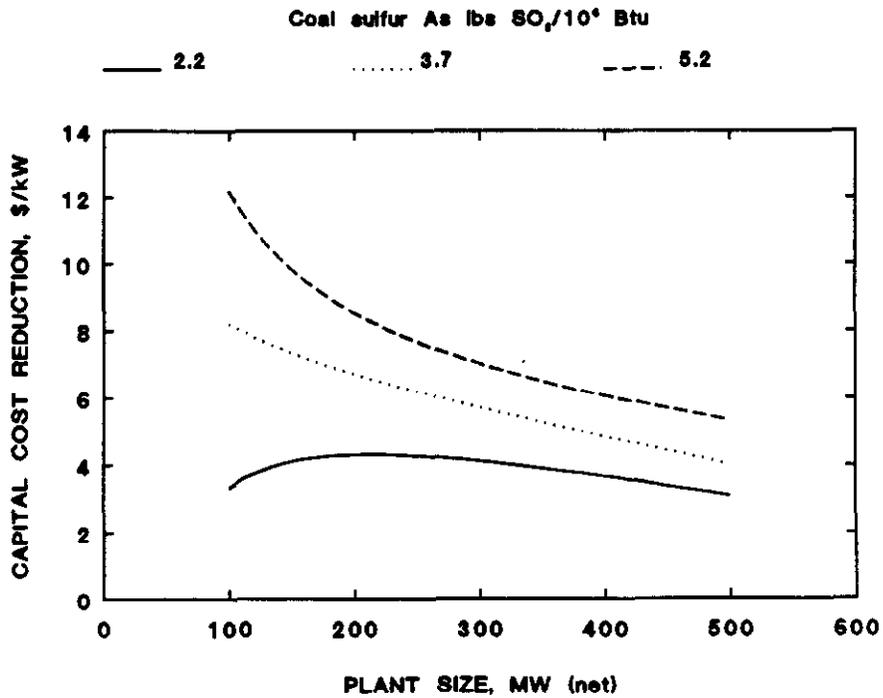


Figure 83. Coolside process capital cost savings for reducing on-site hydrated lime storage from 30 days to 15 days supply.

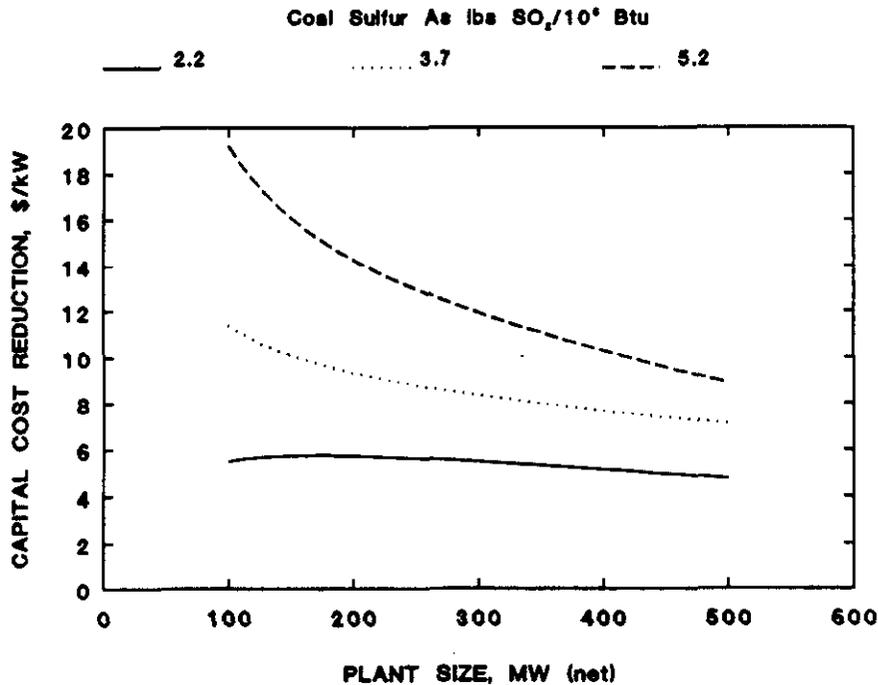


Figure 84. Coalside process capital cost savings for reducing on-site hydrated lime storage from 30 days to 7 days supply.

storage case, cost reductions are \$3-\$12/kW depending upon plant size and coal sulfur level. Reducing the on-site sorbent storage to 7 days supply increases the savings to \$5-\$19/kW. For the optimized Coalside process design, a 15 day on-site sorbent storage is assumed to be adequate.

The shape of the cost saving curves for the low-sulfur coal cases (shown in Figures 83 and 84) is due to modeling procedures which optimize selection of silo type (stave versus reinforced concrete) and number of silos required.

#### Flue Gas Humidifier

The effects of isolation dampers, spare air compressors, and two humidifier versus one humidifier designs were evaluated. Since the cost of these items is dependent only on boiler size, coal sulfur content does not affect the capital cost.

### Damper Costs--

Many utility boilers including small 100 MW units are equipped with dual air heaters and ducts leading to the particulate collection equipment. For such systems, a retrofit Coolside process design might include two separate humidifiers as shown in Figure 85. The design would provide three isolation dampers and one flow control damper for each humidifier. This design allows the humidifiers to be put into or taken out of service without affecting the operation of the boiler. The inclusion of flow control dampers helps to prevent boiler flue gas system pressure swings and upsets when the humidifiers are put into or taken out of service. For a mature Nth design commercial plant, the humidifier operability and reliability should be quite high and there should be no need for dampers. Such a system would be similar to many of the European wet scrubber installations where absorber bypass is not used.

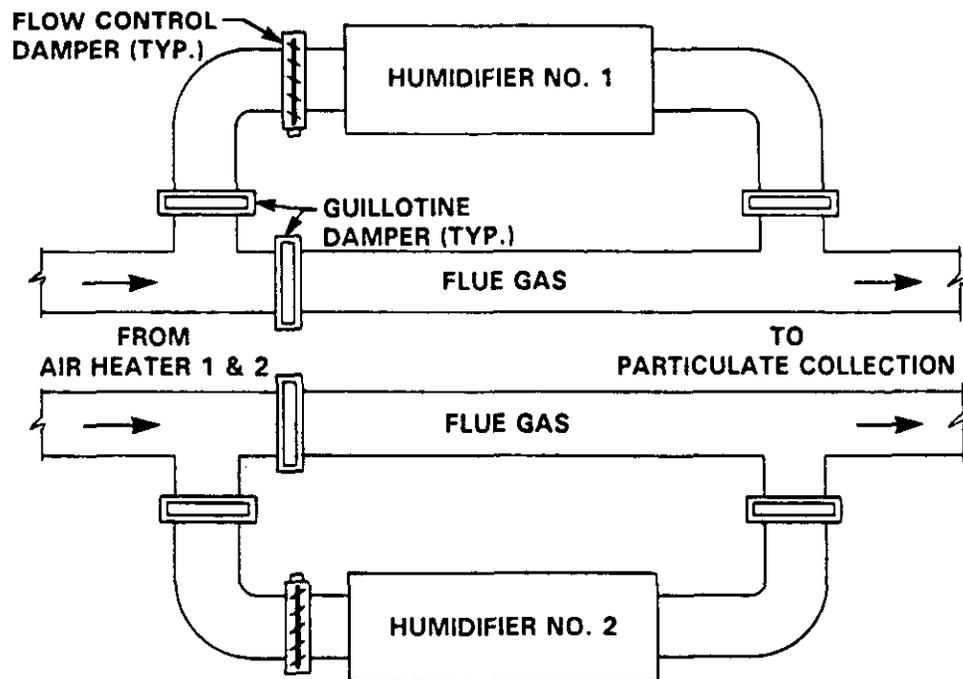


Figure 85. Coolside process humidifier layout for boiler systems with dual flue gas ducts.

As shown in Figure 86 eliminating flue gas dampers can reduce the Coolside process capital costs from \$9.50-\$19.00/kW depending upon the plant size. Dampers are assumed to be unnecessary in an optimized Coolside process. This assumption is also applied to the LSFO FGD economics which are presented later.

#### Single Versus Dual Humidifiers--

As discussed in the preceding section, many utility boilers are equipped with two air preheaters--each with a separate duct to the particulate collection equipment. The effect of combining the flue gases from separate air preheaters into a single humidifier versus the use of a dual humidifier design is presented in Figure 87. The economies of scale afforded by a single large humidifier module reduce the capital costs from \$5-\$23/kW. The largest costs savings are achieved for the smaller plants. A single humidifier is assumed for optimized process designs.

#### Air Compressor Sparing--

The effect of air compressor sparing is shown in Figure 88. The figure shows capital costs savings which can accrue by eliminating the spare compressor. In this case, it is assumed that the Coolside process would be designed with three 50% compressors so that system capacity would be maintained if one compressor failed. Because of high reliability, compressors are not spared in many industries. Therefore, compressor sparing was not required in the optimized Coolside process design. By not providing compressor spares, Coolside process capital cost reductions of \$5-\$14/kW are obtained.

#### Waste Silo Capacity--

Coolside process wastes can be stored in the boiler plant ash silo. However, the silo capacity must be checked in any retrofitted system because of the higher volume of waste solids generated by the Coolside process. At the Edgewater Station, the existing ash silo capacity was sufficient to meet the intermediate storage needs. For the optimized Coolside process design, no additional waste silo storage was considered. However, estimates of the capital costs which would be incurred for undersized systems are provided in Figure 89. The costs are for providing four additional days of waste solids storage with the plant operating at 100 percent of capacity. The additional waste storage capacity is adequate

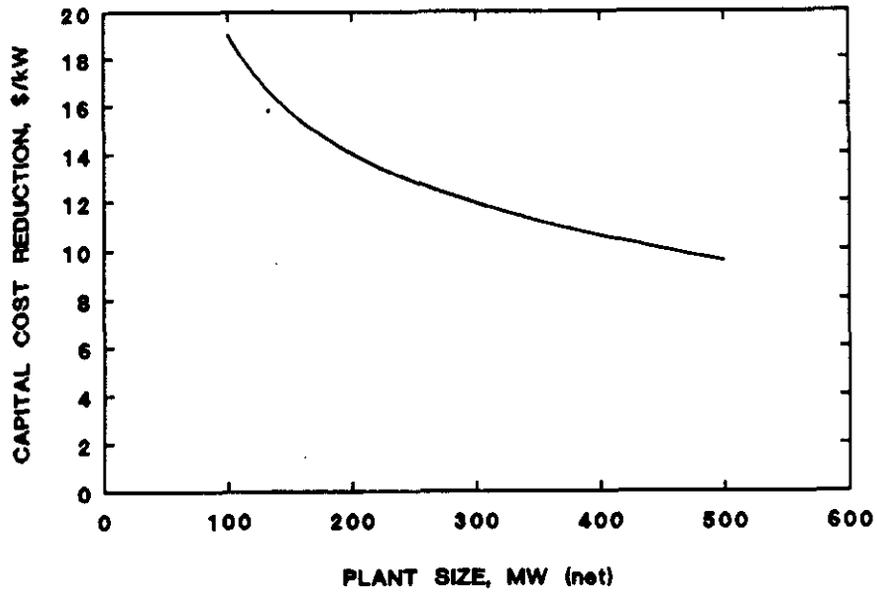


Figure 86. Capital cost reduction for eliminating flue gas duct dampers.

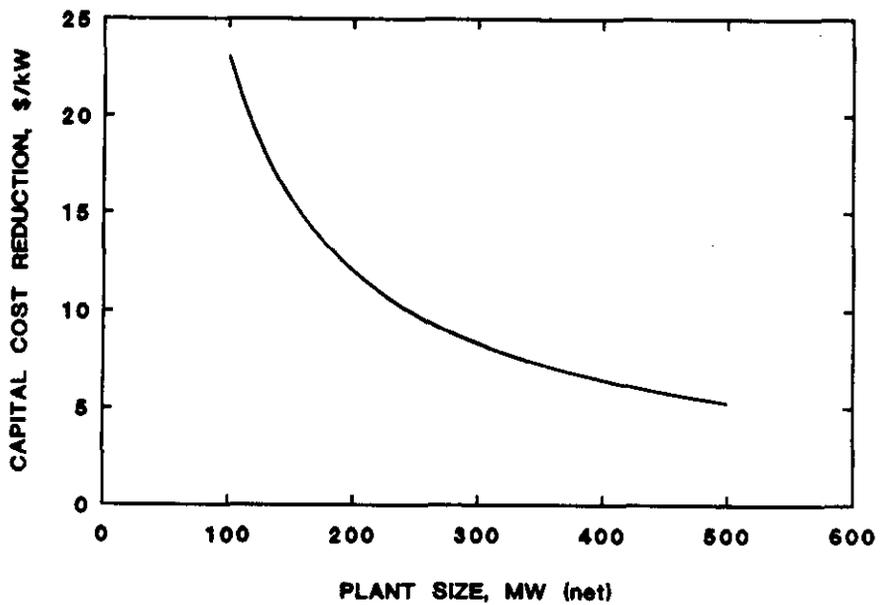


Figure 87. Coolside process capital cost reduction achieved by use of single large module humidifier design.

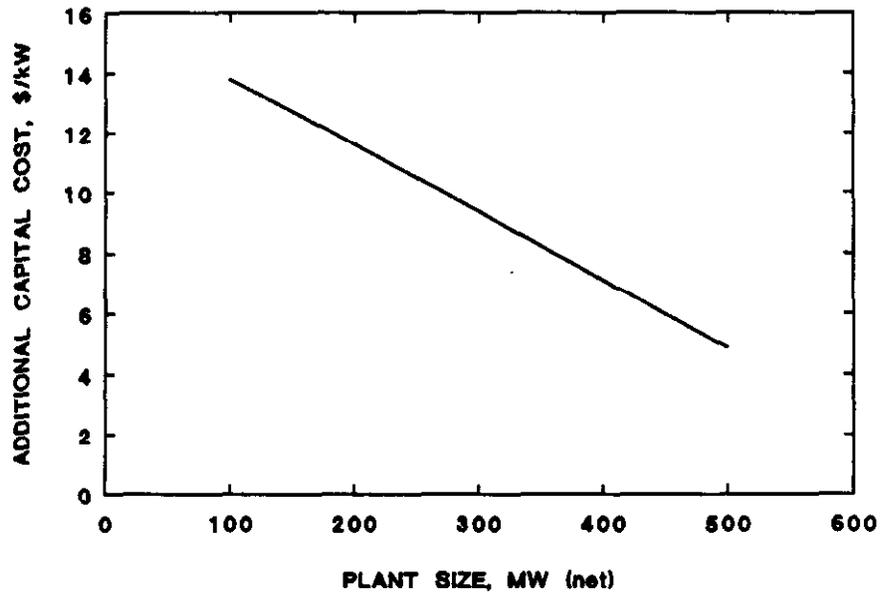


Figure 88. Coolside capital costs associated with providing 50 percent air compressor sparing.

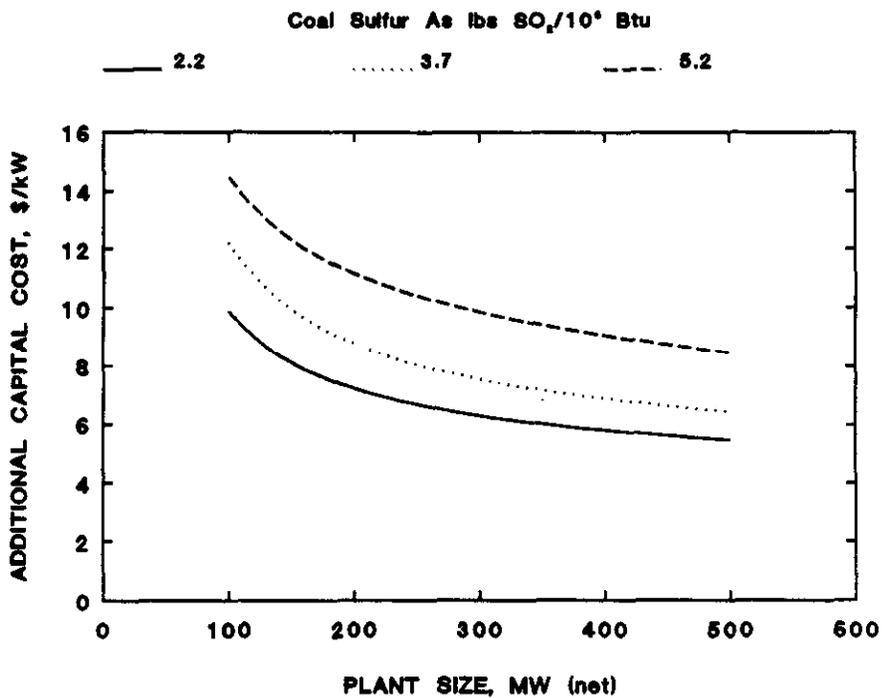


Figure 89. Coolside capital costs for providing four days of additional waste solids storage.

to allow plant operations over extended holiday periods when waste removal to a landfill may not be available. The costs include the costs for the silo or silos and dustless unloading equipment. For the 2.5 weight percent sulfur (3.7 lbs  $\text{SO}_2/10^6$  Btu) coal, providing the additional storage would increase the capital costs by about \$6.50-\$12.00/kW, again depending upon the plant size.

The effect of varying the waste disposal charge on levelized costs is presented in Figure 90. The base cost is \$8.60/ton of dry wastes with a \$7.00/ton fly ash credit. Because fly ash is collected with the Coolside process waste solids, a fly ash credit is assigned. This is justified since a utility will need to dispose of fly ash whether or not the boiler has an FGD system. No fly ash credit is assigned to the LSFO FGD because the fly ash is removed prior to the FGD and is therefore not contained in the waste gypsum.

Reducing the waste disposal costs to \$7.00/ton (equivalent to the assumed fly ash disposal cost) would reduce levelized cost by about \$9-\$11/ton of  $\text{SO}_2$  removed depending upon the coal sulfur content. Nearly doubling the waste disposal charge to \$15.00/ton increases the levelized costs by about \$36-\$43/ton of  $\text{SO}_2$  removed. For this comparison, the fly ash disposal charge was held constant (\$7.00/ton). The differential costs for the medium- and high-sulfur coals are nearly identical. This is because the total weight of waste including fly ash per unit weight of  $\text{SO}_2$  removed is nearly identical due to the interaction between ESP efficiency (which limits the amount of recycle possible) and increasing sorbent feed rate required to maintain the  $\text{SO}_2$  removal.

To summarize, the optimized Coolside process design is based on providing no spare air compressors, no isolation dampers, and a single humidifier for all reference plants. For the economics which follow, it has been assumed that no additional ESP area or waste solids storage would be required. Minor ESP upgrading of the rapper system is assumed to be required. Costs are provided for replacing low-intensity rappers with high-intensity tumble hammer rappers.

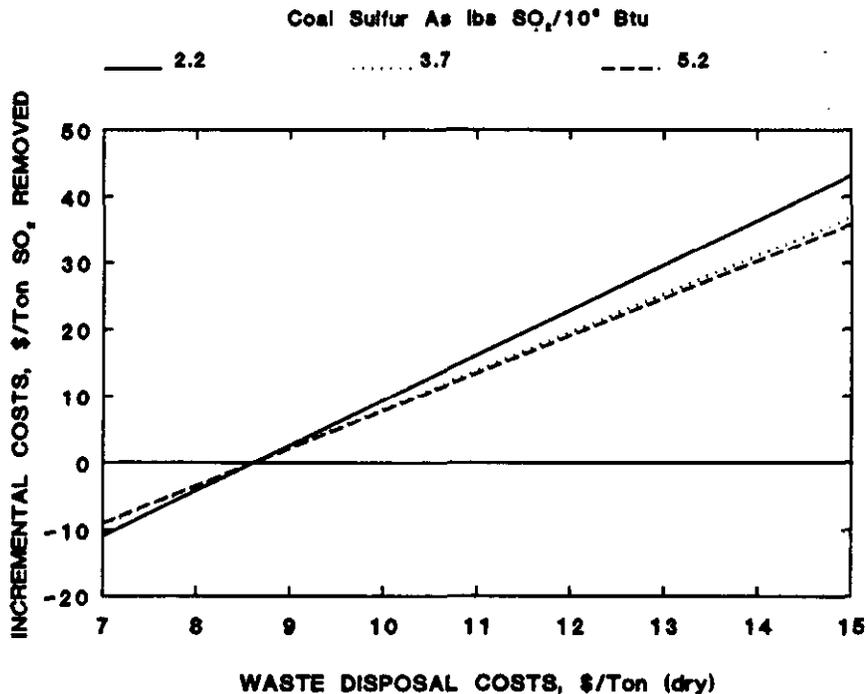


Figure 90. Effect of waste disposal charge on levelized Coolside process costs.

### FGD Design Basis

The wet limestone forced oxidation (LSFO) FGD process was selected for economic comparison with the Coolside process. The process is shown in Figure 91. Limestone is delivered by rail to the site and then transferred to a 30-day covered storage pile. The limestone is then fed to a day bin and, as required, to the grinding equipment for pulverization to 90% minus 325 mesh in a wet ball mill. The limestone slurry produced is metered as needed to the SO<sub>2</sub> absorber. The absorber is a single module (rubber-lined, carbon steel, open spray tower) which treats the flue gas from the boiler. Boiler flue gas passes through booster fans and then enters the absorber where the gas is contacted with a limestone slurry. The SO<sub>2</sub> reacts with the slurry forming calcium sulfite, CaSO<sub>3</sub>•1/2H<sub>2</sub>O, and calcium sulfate (gypsum), CaSO<sub>4</sub>•2H<sub>2</sub>O. The reacted slurry collects in the absorber sump. Air is blown into the absorber sump at a stoichiometric ratio of 1.5 mol O<sub>2</sub>/mol SO<sub>2</sub> to convert all the sulfite to

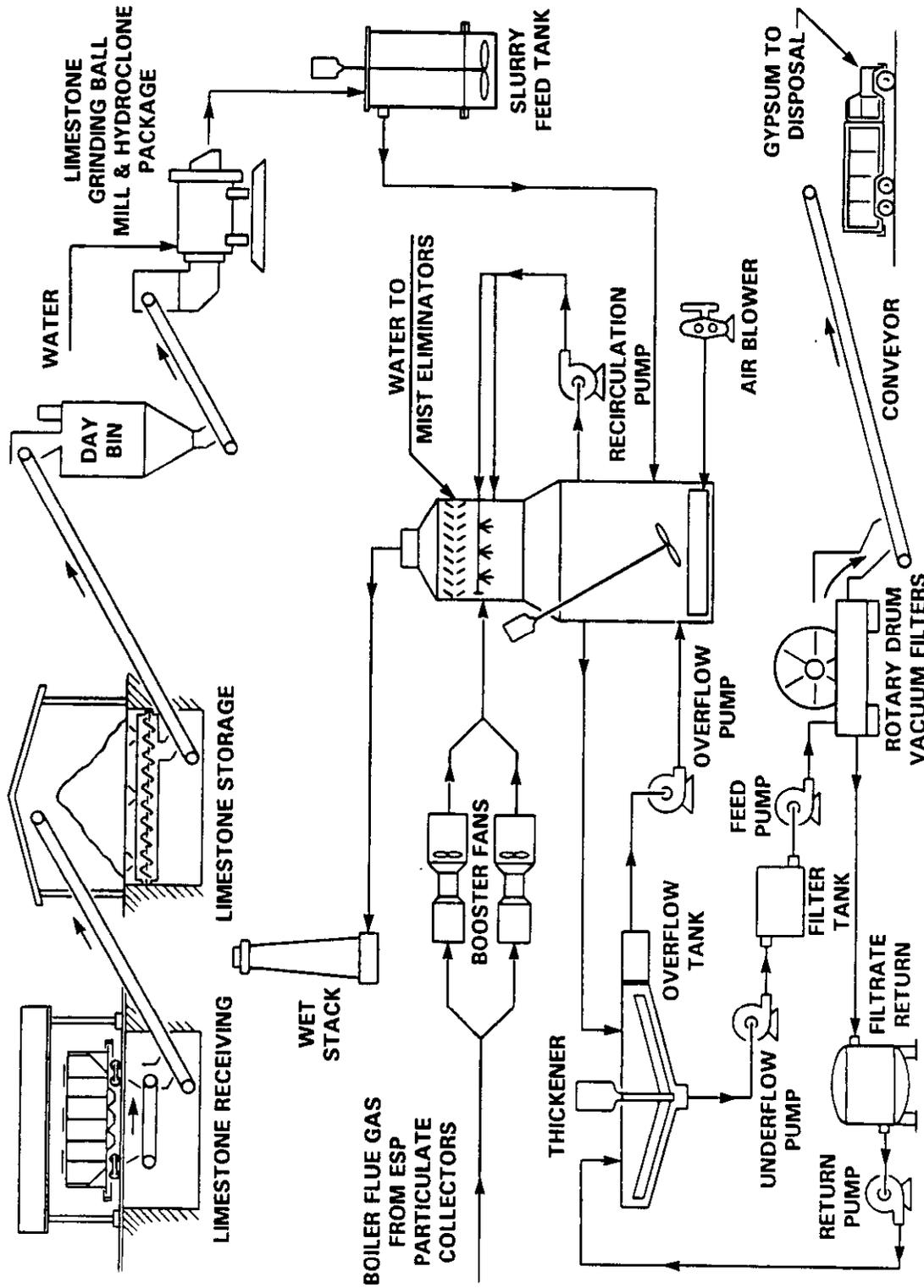


Figure 91. Wet limestone forced oxidation (LSFO) process.

sulfate. Large slurry pumps take suction from the absorber sump and recycle the slurry back through nozzle equipped spray headers in the absorber. An absorber sump bleed stream is sent to a thickener for partial dewatering. The thickener overflow returns to the absorber while the underflow is pumped to vacuum rotary drum filters. Gypsum is dewatered to 80% solids and disposed in a landfill. The clean flue gas is exhausted through a new wet stack.

For the process design, the limestone grinding mills, recycle and slurry transfer pumps, the oxidation air blowers, the waste solids filters, and the gypsum product conveyor are spared. The booster fans, thickener, and absorber are not spared. Like the Coolside process design, there is no absorber bypass and no flue gas dampers. The design SO<sub>2</sub> removal is 95%. The waste disposal capital cost is assumed to be recovered in the fixed waste disposal charge rate, \$/ton disposal. The economics for the LSFO FGD are calculated in the same fashion as described in the Economic Premises Section. A typical computer model output is shown in Appendix D for the reader's reference. Yearly rates for consumables and utilities are presented for reference plants in Appendix E.

## OPTIMIZED COOLSIDE PROCESS AND WET LIMESTONE FGD COST COMPARISON

### Capital Cost Comparison

The published capital cost estimates for sorbent injection and wet limestone scrubbing vary over a wide range, depending on the process and economic assumptions used in the studies. This study used an internally consistent set of process and economic assumptions in developing the capital cost estimates. Therefore, the relative comparison of Coolside and wet FGD economics is considered to be generally valid. Site-specific factors will determine the absolute costs for actual applications.

The optimized Coolside capital costs and optimized LSFO FGD capital costs are plotted in Figures 92, 93, and 94 for the design coals. In all cases, LSFO FGD capital costs are much higher than the Coolside process capital costs. The LSFO FGD capital costs are 2.2 to 2.5 times the Coolside costs. In cases where high SO<sub>2</sub> removals are not required, lower capital cost favors the installation of the Coolside process.

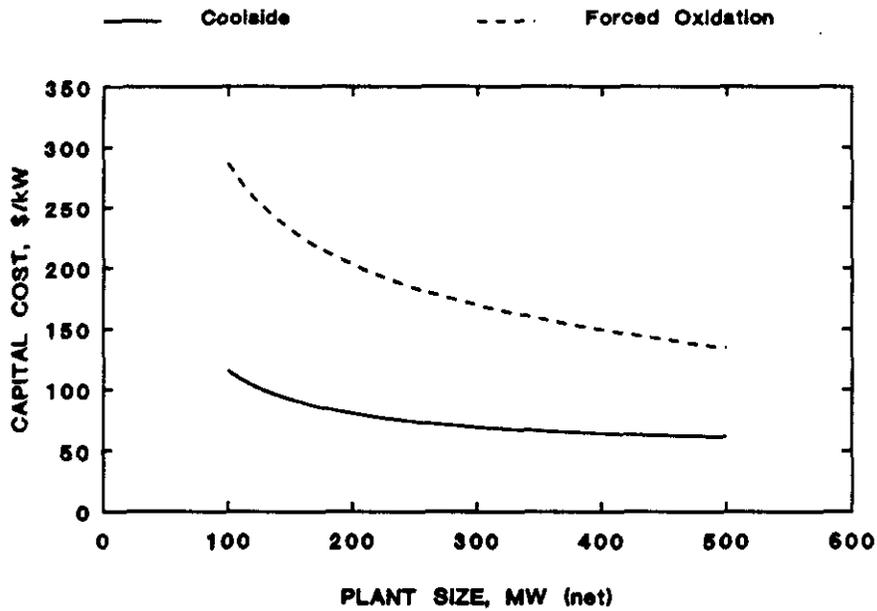


Figure 92. Capital cost comparison for Coolside versus LSF0 (coal sulfur equivalent to 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

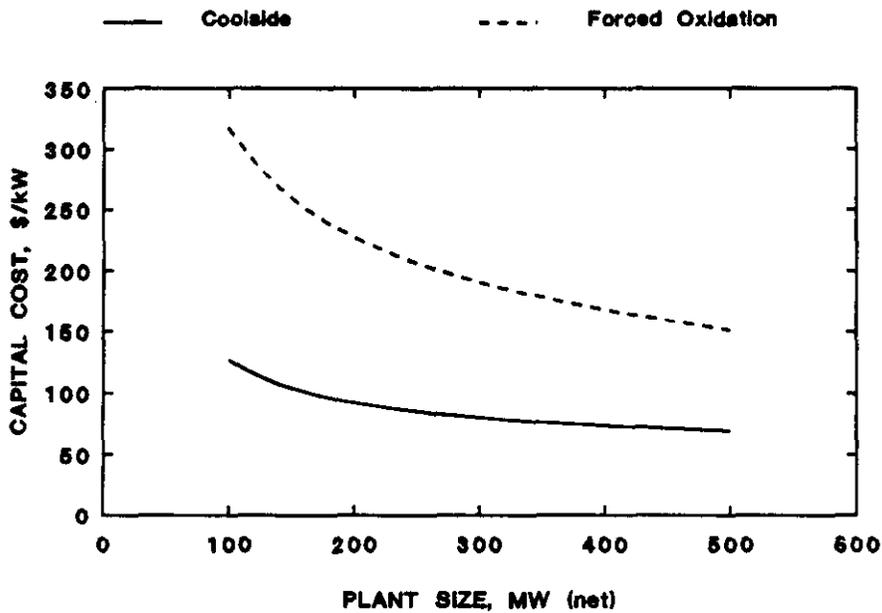


Figure 93. Capital cost comparison for Coolside versus LSF0 (coal sulfur equivalent to 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

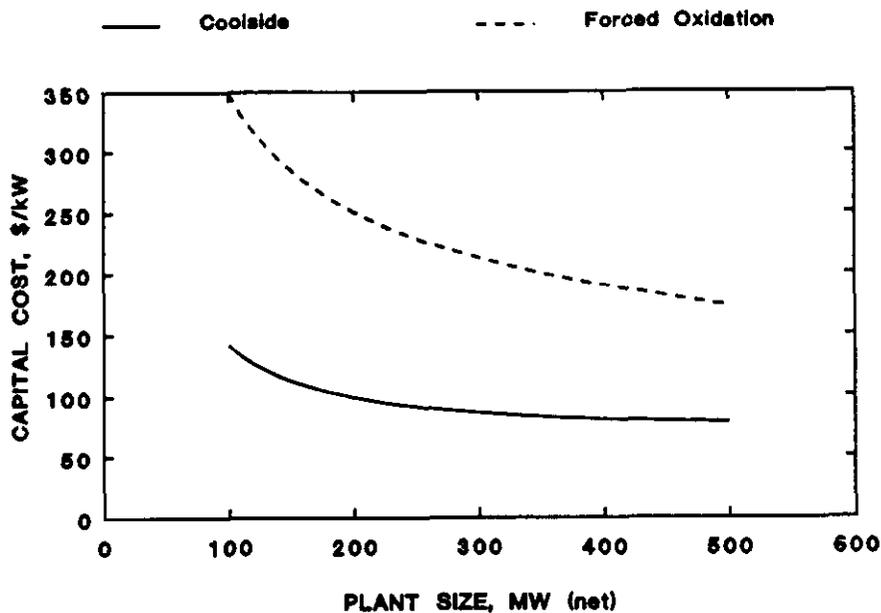


Figure 94. Capital cost comparison for Coolside versus LSF0 (coal sulfur equivalent to 5.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

### Levelized Cost Comparison

The total levelized annual revenue requirements (see Table 16 for economic basis) in constant dollars (mid 1990) for both the Coolside and LSF0 processes are presented in Figures 95, 96, and 97. The base plant capacity factor is 65%. For the 1.5 weight percent sulfur coal case (see Figure 95), the Coolside process is economically competitive with LSF0 FGD up to a crossover point plant size of about 350 MW<sub>e</sub>. For the 2.5 weight percent sulfur coal case (Figure 96), the Coolside process is economically competitive up to about 130 MW<sub>e</sub> plant capacities. The LSF0 FGD is preferred for the 3.5 weight percent sulfur coal (Figure 97) over the size range of reference plants studied.

For the base case economics (65% plant capacity factor) discussed above, the Coolside process window of applicability is for plants up to 350 MW<sub>e</sub> for 1.5

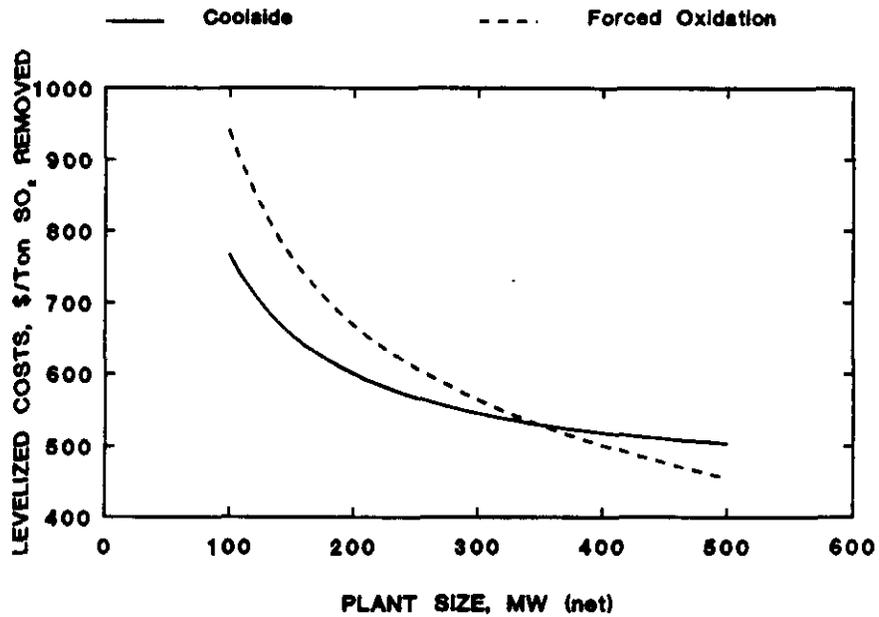


Figure 95. Levelized cost comparison for Coolside versus LSF0 (65 percent plant capacity factor, coal sulfur equivalent to 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

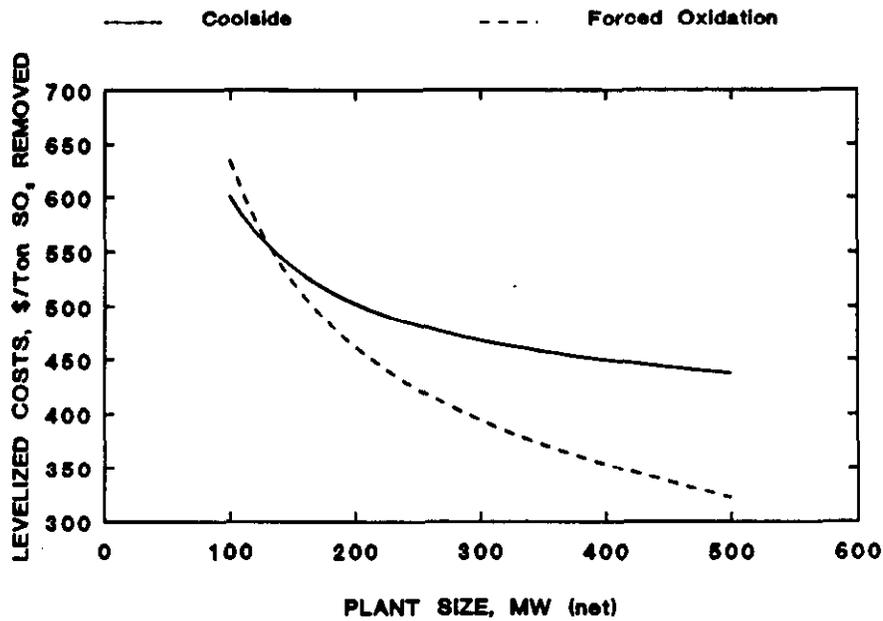


Figure 96. Levelized cost comparison for Coolside versus LSF0 (65 percent plant capacity factor, coal sulfur equivalent to 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

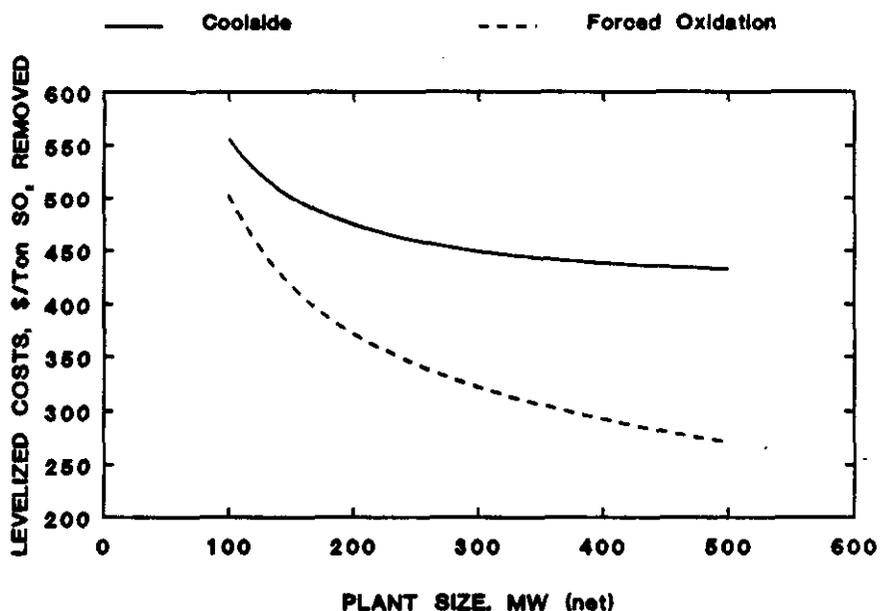


Figure 97. Levelized cost comparison for Coalside versus LSF0 (65 percent plant capacity factor, coal sulfur equivalent to 5.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

weight percent sulfur coal and up to 130 MW<sub>e</sub> for 2.5 weight percent sulfur coal. Factors which can alter this analysis are plant capacity factor, required SO<sub>2</sub> removal, difficulty of retrofit, and remaining plant life. Figures 98, 99, and 100 show the effect of the plant capacity factor on economics. With lower sulfur coal (1.5 weight percent sulfur, 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu) and a plant capacity factor of 50%, the Coalside process economic competitiveness with LSF0 extends up to a plant size of 475 MW<sub>e</sub>. For the medium-sulfur coal (2.5 weight percent sulfur, 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu), the range of applicability now extends to 220 MW<sub>e</sub> at 50% capacity factor and to 320 MW<sub>e</sub> at 40% capacity factor (see Figure 99). For the high-sulfur coal (3.5 weight percent sulfur, 5.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu) and a 40% capacity factor, the Coalside process is competitive with LSF0 at the 170 MW<sub>e</sub> scale assuming the 70% SO<sub>2</sub> removal efficiency is a justifiable emission control option for the plant site. These results clearly show that lower capacity factor units favor low capital cost, higher operating cost processes such as the Coalside process, and that the range of process applicability for the Coalside process significantly expands as capacity factor is lowered.

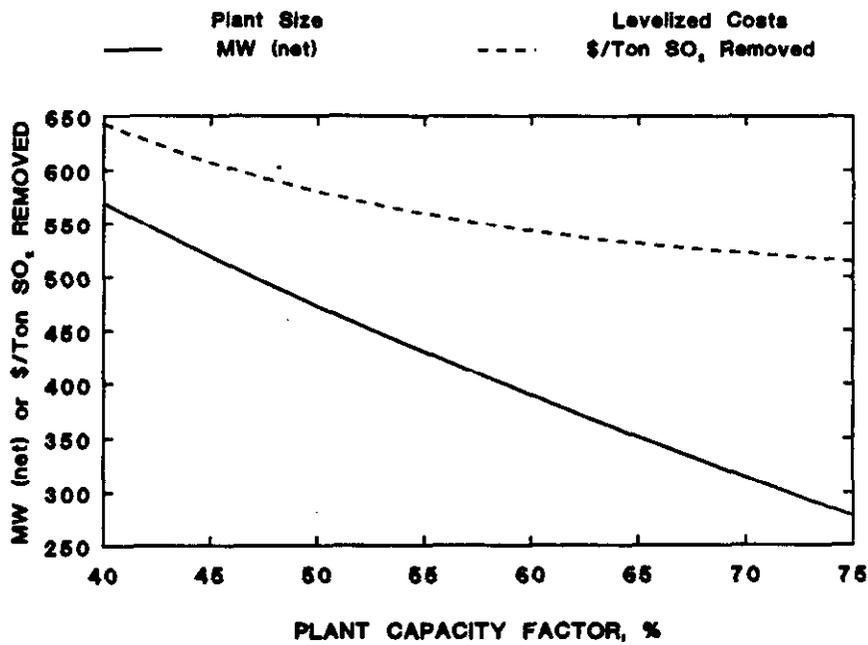


Figure 98. Effect of plant capacity factor on the plant size at the economic crossover point between Coalside and LSFO processes and levelized costs (coal sulfur equivalent to 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

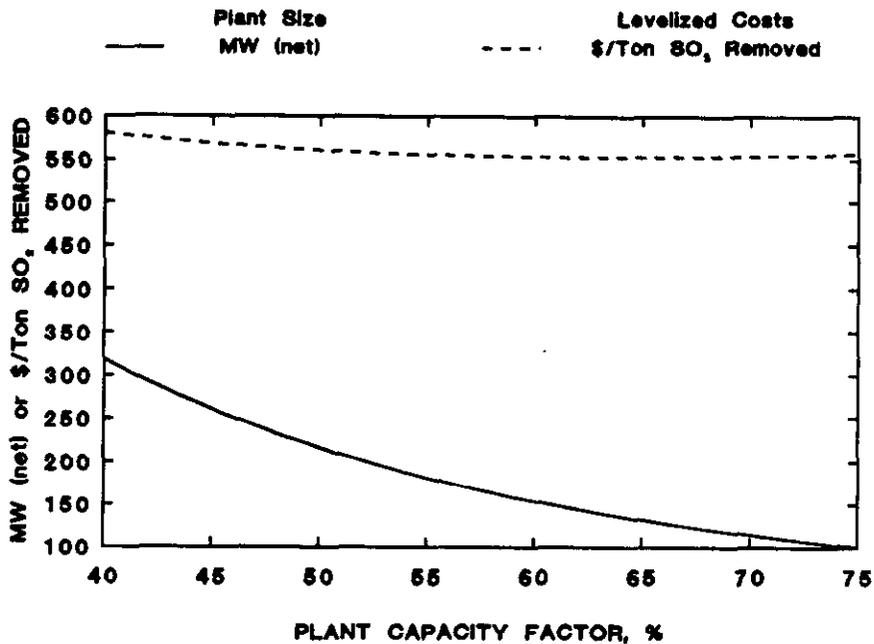


Figure 99. Effect of plant capacity factor on the plant size at the economic crossover point between Coalside and LSFO processes and levelized costs (coal sulfur equivalent to 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

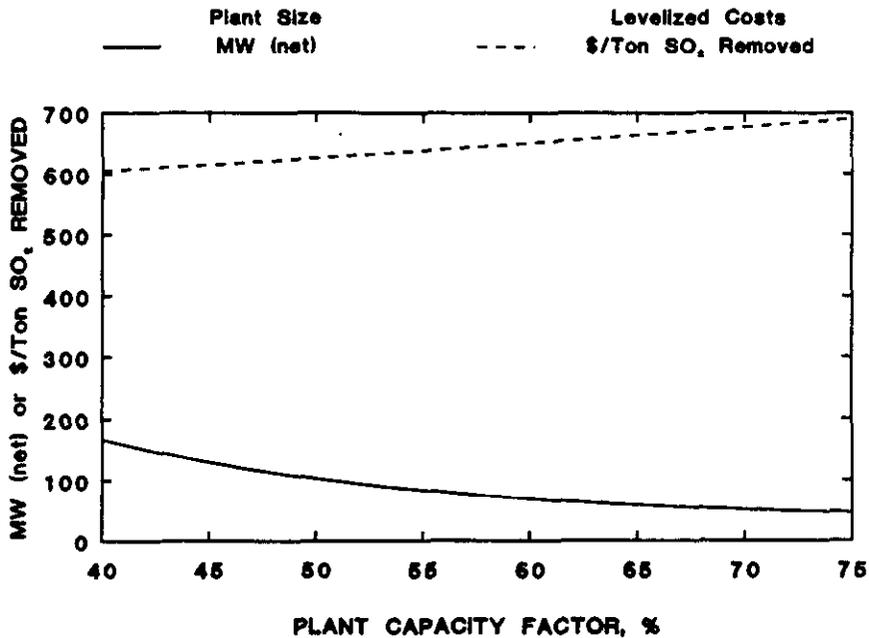


Figure 100. Effect of plant capacity factor on the plant size at the economic crossover point between Coalside and LSFO processes and levelized costs (coal sulfur equivalent to 5.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

Throughout this study, the LSFO FGD SO<sub>2</sub> removal is assumed to be 95% on an annual basis while the Coalside process SO<sub>2</sub> removal is assumed to be 70%. In general, the Coalside process becomes economically more favored relative to LSFO FGD as SO<sub>2</sub> removal requirements are reduced. For example as shown in Table 17, for an optimized 250 MW<sub>e</sub> (net) LSFO FGD treating the flue gas from the combustion of a 2.5% sulfur coal, lowering the SO<sub>2</sub> removal requirement from 95% to 70%, then to 50%, reduces the capital costs from \$207/kW (95% removal) to \$189/kW (70% removal), to \$167/kW (50% removal), respectively. However, because of high capital charges, the cost per ton of SO<sub>2</sub> removed increases from \$426/ton (95% removal) to \$512/ton (70% removal), then to \$630/ton (50% removal). Similarly, lowering the required Coalside SO<sub>2</sub> removal from 70% to 50% reduces the capital costs from \$87/kW to \$74/kW. The Coalside levelized costs per ton of SO<sub>2</sub> removed decrease slightly from \$488/ton (at 70% removal) to \$481/ton (at 50% removal).

To achieve the Clean Air Act Amendments of 1990 Phase II base year emission limit 1.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu, the Coalside process operating at 70% SO<sub>2</sub> removal could treat the flue gas from the combustion of coal containing up to 2.5 weight percent sulfur. As a result, for lower sulfur coals, SO<sub>2</sub> control credits might

be available for sale or use at other installations. Lower SO<sub>2</sub> removal requirements favor the installation of the Coolside process at sites where cheaper, near compliance, local coal can be used in place of more expensive, out-of-state, compliance coal.

Table 17  
EFFECT OF SO<sub>2</sub> REMOVAL LEVEL ON COSTS  
FOR NOMINAL 250 MW<sub>e</sub> (NET) PLANT SIZE\*

Process	SO <sub>2</sub> Removal %	Capital Cost \$/kW (net)	Levelized Costs \$/ton SO <sub>2</sub> Removed
LSFO	95	207	426
LSFO	70**	189	512
LSFO	50**	167	630
Coolside	70	87	488
Coolside	50	74	481

\*Optimized processes, 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu coal (2.5 wt % S), 65% plant capacity factor.

\*\*95% SO<sub>2</sub> removal absorbers with flue gas bypass.

For the economic comparison in this report, the average retrofit factor is about 1.3 (moderately difficult retrofit) for both the wet scrubber and Coolside processes. For most situations, the Coolside process will have a smaller footprint than the LSFO FGD. Relatively little equipment is associated with the Coolside process, and the flue gas humidifier will normally have one-half the cross sectional area of a wet scrubber handling the same gas flow. Therefore, the Coolside process may be easier to retrofit in many cases than a wet scrubber. The resulting difference in retrofit factor would increase the capital cost of the wet scrubber relative to the Coolside process and further increase the attractiveness of the Coolside process. A site-specific, retrofit analysis should be conducted at each site to determine the correct retrofit factor for all options to be considered.

Another factor which can affect the economic analysis is remaining plant life. The base case assumption is 20 year plant life for retrofit installations. At some older plant sites, a 10 year project life may be appropriate. For two

options which have the same life cycle levelized costs, reducing the plant life will favor installation of the lowest capital cost option because the capital charges increase proportionally for both options while the operating expenses remain constant. Lower plant life favors the lower capital cost Coolside process.

### Short-Term Analysis

Short-term economic analysis is used by many utilities to determine if the long range benefits of an option show an economic advantage in the short-term when unforeseen factors such as changes in regulatory constraints or technical obsolescence are less likely to occur. Typically short-term analysis considers only the first 10 years of the project life.

The results of a short-term (10 year) cost analysis are presented in Figures 101, 102, and 103 for low-, medium-, and high-sulfur coal cases, respectively. The analysis shows that the window of applicability for the Coolside process is extended to larger plants by about 50 MW<sub>e</sub> in the low-sulfur coal case and by about 30 MW<sub>e</sub> in the medium-sulfur coal case. For high-sulfur coal, LSFO is still favored over the Coolside process. The economic assumptions in this analysis are the same as the 20 year levelized analysis. The only change is the use of the constant dollar levelizing factor (shown in Table 16) for short-term capital carrying charges.

### Effect of Coal Quality On Costs

The economics presented in this report are based on using a coal with an as-received (AR) higher heating value (HHV) 13,400 Btu/lb and an ash content 7.6 wt % (AR). For a given coal sulfur content, the potential SO<sub>2</sub> emissions (lb SO<sub>2</sub>/10<sup>6</sup> Btu) will be lower for a coal with a greater HHV. This reduces reagent requirements for both the Coolside and LSFO processes. Lower coal ash contents improve sorbent utilization for the Coolside process because more solids can be recycled from the particulate collection device.

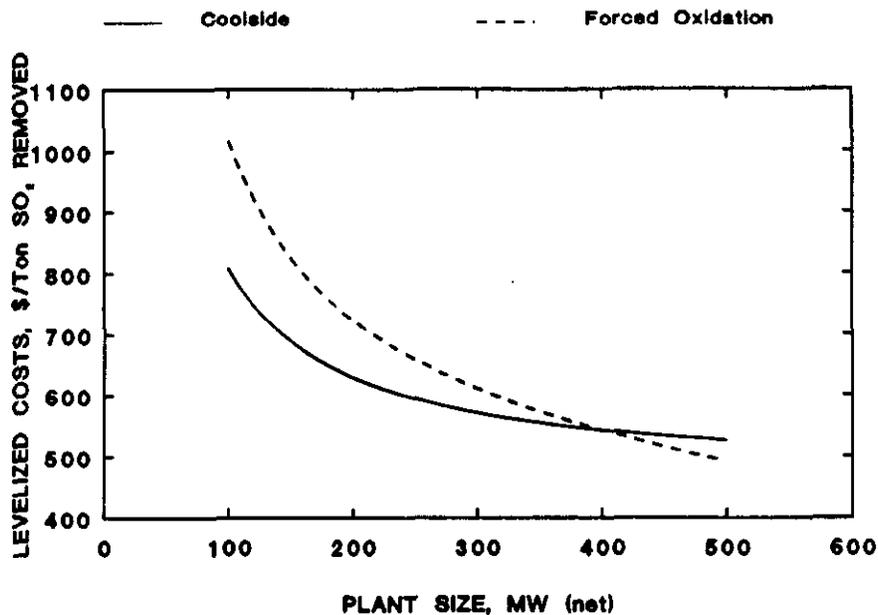


Figure 101. Short-term (10 year) cost comparison for Coolside process versus LSF0 (65 percent plant capacity factor, coal sulfur equivalent to 2.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

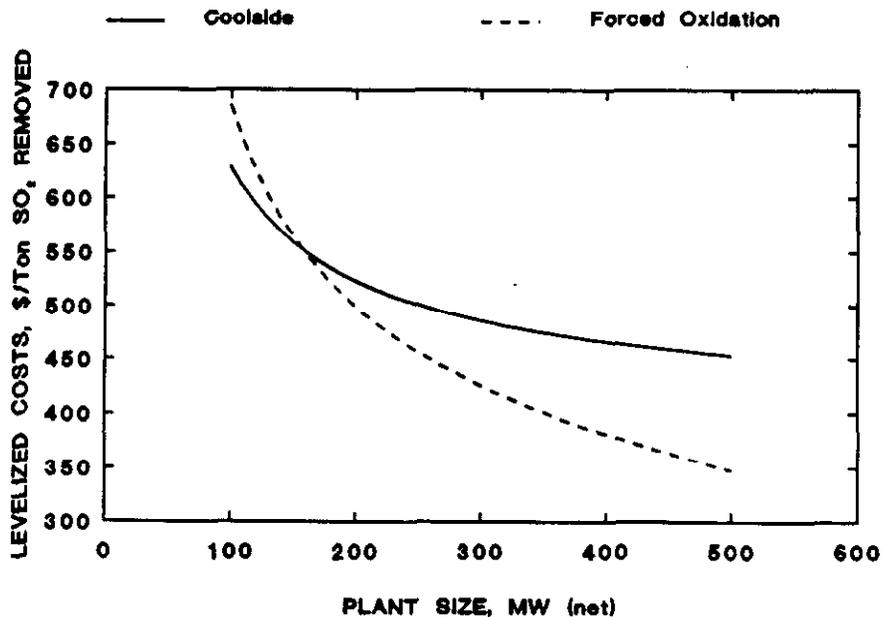


Figure 102. Short-term (10 year) cost comparison for Coolside process versus LSF0 (65 percent plant capacity factor, coal sulfur equivalent to 3.7 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

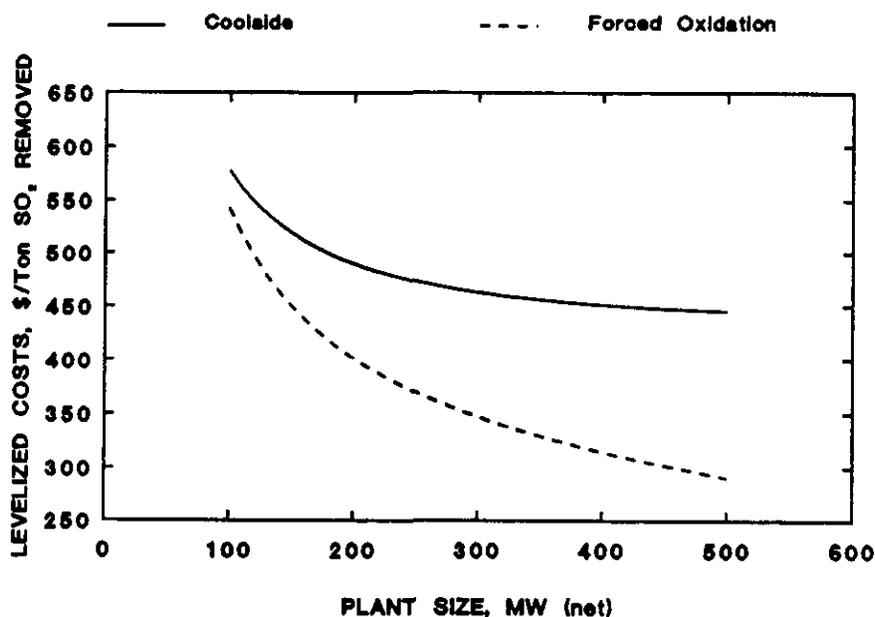


Figure 103. Short-term (10 year) cost comparison for Coolside process versus LSFO (65 percent plant capacity factor, coal sulfur equivalent to 5.2 lb SO<sub>2</sub>/10<sup>6</sup> Btu).

The effect of using a lower quality coal on process economics is presented in Table 18. Comparisons are presented for 100 MW<sub>e</sub> (net) and 250 MW<sub>e</sub> (net) nominal plant sizes burning 1.5 wt % sulfur coals. The lower quality coal used in the comparison has an HHV of 11,872 Btu/lb (AR) and an ash content of 10.77 wt % (AR). The coal is very similar to the low-sulfur coal used at Edgewater during the test program.

Decreasing the coal heating value by 11% from 13,400 Btu/lb to 11,872 Btu/lb increases the potential SO<sub>2</sub> emissions by about 13% from 2.24 lb SO<sub>2</sub>/10<sup>6</sup> Btu to 2.53 lb SO<sub>2</sub>/10<sup>6</sup> Btu. As shown in Table 18, the effect of this change on the Coolside process capital and levelized costs is relatively minor. The 13% increase in the amount of SO<sub>2</sub> removed for the lower quality coal only increases the capital costs for the 100 MW<sub>e</sub> and 250 MW<sub>e</sub> plant sizes by about 3% and 5%, respectively. The levelized costs in terms of \$/ton of SO<sub>2</sub> removed do not change much either because more SO<sub>2</sub> must be removed for the lower quality coal cases. For the 100 MW<sub>e</sub> plant size, the levelized costs decrease by about 2.6% and are essentially neutral for the 250 MW<sub>e</sub> plant size.

Actual out-of-pocket operating expenses are, however, greater for the lower quality coal mainly because sorbent usage increases due to the higher amount of sulfur which must be removed and a higher required fresh Ca/S ratio. The fresh Ca/S ratio increases from 1.33 mol/mol for the original coal to 1.52 mol/mol for the lower quality coal. Increased coal usage and fly ash levels for the lower quality coal result in a reduction in the recycle solids rate to limit the inlet solids loadings to the particulate collector. This increases the required fresh sorbent makeup. In general, the use of higher quality coals with higher heating values and lower ash contents reduces SO<sub>2</sub> control system operating costs.

Table 18  
EFFECT OF COAL QUALITY ON COSTS  
65 PERCENT PLANT CAPACITY FACTOR, 70 PERCENT SO<sub>2</sub> REMOVAL

Nominal Plant Size MW <sub>e</sub> (net)	Coal Sulfur Wt %	HHV Btu/lb (AR)	SO <sub>2</sub> Potential lb/10 <sup>6</sup> Btu	Capital Cost \$/kW (net)	Levelized Cost \$/ton SO <sub>2</sub> Removed
100	1.5	13400	2.24	116.1	772
100	1.5	11872	2.53	119.7	752
250	1.5	13400	2.24	73.7	567
250	1.5	11872	2.53	77.4	569

## SECTION 6

### PILOT-SCALE (0.1 MWe) SUPPORT TESTS

#### PILOT TEST PROGRAM OBJECTIVES

The objectives of the 0.1 MWe pilot-scale support program for the Edgewater demonstration were: (1) to identify sorbents for use in the demonstration, and (2) to develop process performance data with and without sorbent recycle over a wide range of process conditions, including Edgewater site-specific conditions. The pilot plant data provided a basis for the demonstration program design and data evaluation. Additionally, the pilot plant data provide a basis for future evaluation of the Coolside process.

The pilot support program consisted of sorbent evaluation, once-through simulation, and recycle simulation tests. The sorbent evaluation involved pilot testing of the desulfurization performance of twelve commercially available candidate sorbents (ten calcitic and two dolomitic hydrated limes) and physical/chemical characterization of these sorbents. The test results were used to select two hydrated limes for use at Edgewater. In the once-through simulation tests, process performance data were generated for the two selected sorbents over a wide range of test conditions using a statistical experimental design. The data were analyzed to develop correlations for SO<sub>2</sub> removal as a function of key process variables. The recycle simulation test generated performance data for the recycle mode of operation under Edgewater site-specific conditions. All the pilot-scale testing was conducted using the Consolidation Coal Company 0.1 MWe Coolside pilot plant.

Results of the sorbent evaluation, once-through simulation and recycle simulation tests were reported previously.<sup>1,4,5,25</sup> The major results of the tests are summarized below.

## DESCRIPTION OF CONSOL COOLSIDE PILOT PLANT

The 0.1 MWe pilot test facility (Figure 104) is designed to study Coolside desulfurization performance over a wide range of process conditions. The unit is described in References 1 and 2. It allows site-specific simulation of flue gas conditions, including gas composition, temperature, solids loading, and residence time. The unit has an 8.3-inch ID x 20-foot long vertical downflow humidifier installed with a commercial two-fluid nozzle. A pulse-jet baghouse is used for particulate removal. Continuous SO<sub>2</sub> and O<sub>2</sub> analyzers are used for measurement of SO<sub>2</sub> removal across the humidifier and across the system (humidifier + baghouse).

The Coolside pilot unit has been used previously for Coolside process variable tests<sup>5</sup> and for studies of Coolside desulfurization mechanisms.<sup>6</sup> In these tests, the operability of the pilot test unit and the reliability of pilot test data were demonstrated successfully. Operating conditions were defined which allowed operation with minimal humidifier wall wetting and minimal wet solids deposition. Solid elemental material balances closed to within  $\pm 10\%$ . Analysis of collected ash material confirmed the SO<sub>2</sub> removals measured by the continuous flue gas analyzers. The desulfurization performance data obtained in the pilot unit are consistent with results of earlier 1 MWe Coolside field tests.<sup>7</sup>

For all the pilot tests conducted in the Edgewater support program, the flue gas flow was fixed at 175 scfm, which gave about a 2-second residence time in the humidifier. A single Spraying Systems J12 atomizer was operated at an air/water ratio of about 50 scf/gal to provide one second or less drying time at 25°F approach to adiabatic saturation.

## SUMMARY OF PILOT TEST RESULTS

### Sorbent Evaluation

Twelve different sorbents were evaluated (Figure 105), including 10 hydrated calcitic limes (designated limes A to J) and two pressure hydrated dolomitic limes (designated limes D-A and D-B). Desulfurization performance varied widely

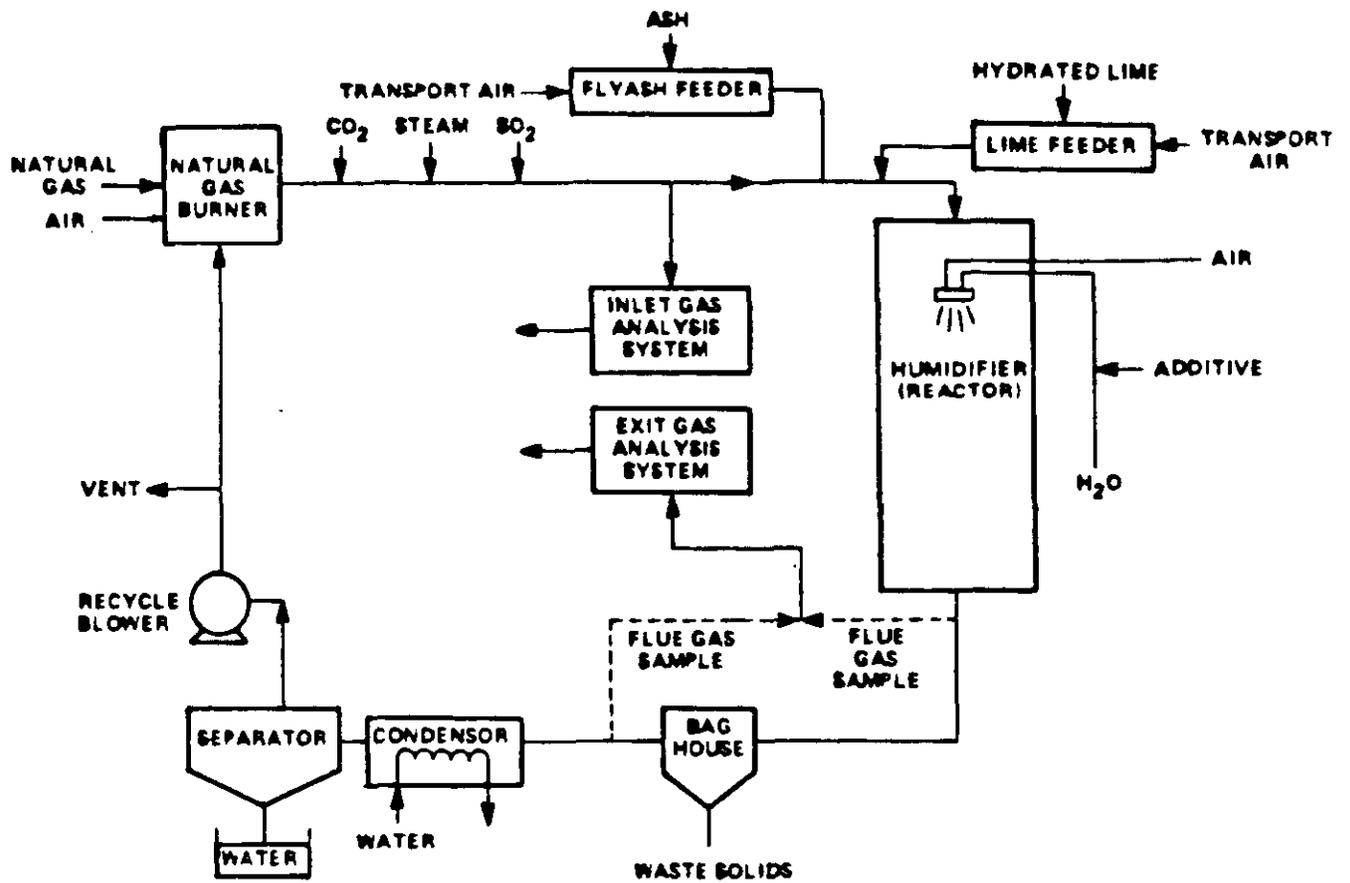


Figure 104. Schematic of pilot unit.

**RELATIVE UTILIZATION EFFICIENCY, (CALCITIC A·100)**

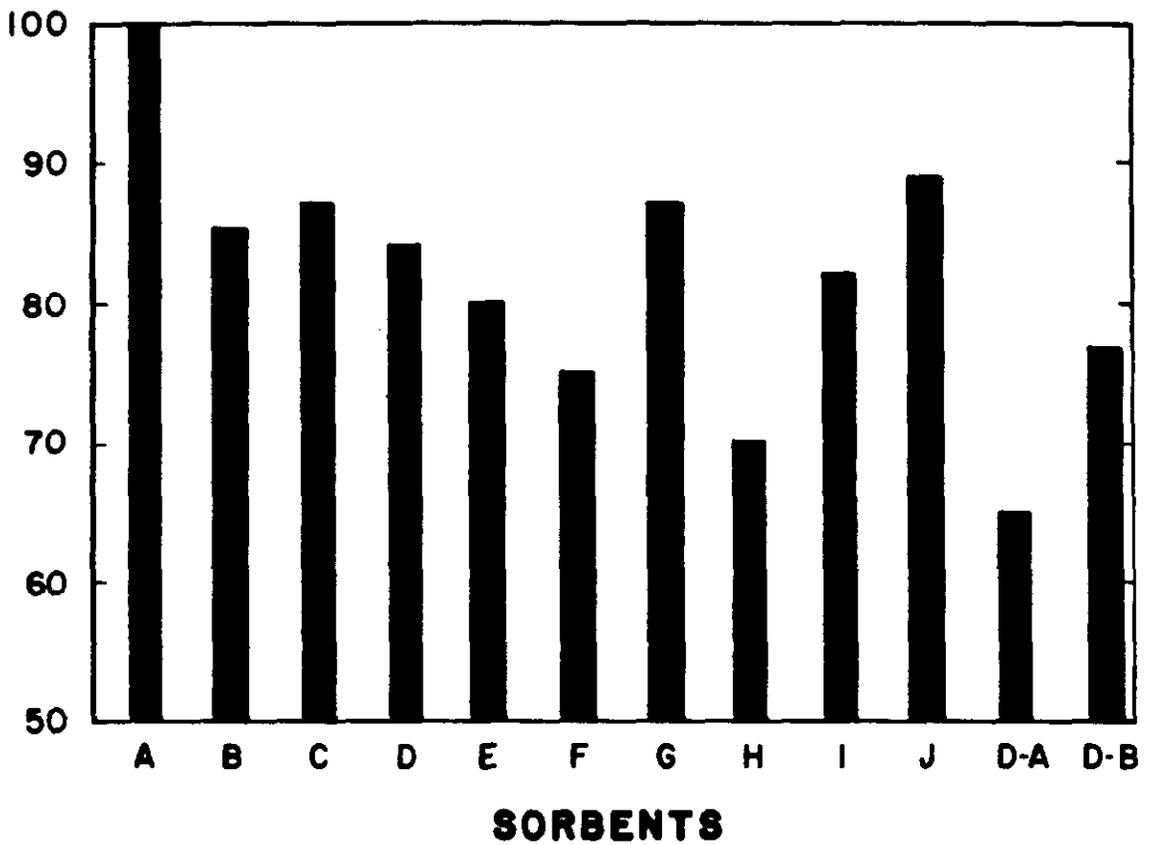


Figure 105. Comparison of relative sorbent efficiency averaged over all Coolside runs. Sorbents A through J are calcitic hydrated limes and D-A and D-B are dolomitic hydrated limes.

among the different sorbents. Based on the test results, two hydrated calcitic limes were selected for testing at Edgewater. The first, hydrated lime A, was the most active sorbent, based on the pilot data. The second, hydrated lime G, had the lowest delivered cost to Edgewater.

Figure 105 compares the average performance of each sorbent relative to the performance of hydrated lime A, based on sorbent utilizations observed in three pilot plant runs at varying process conditions. The relative sorbent efficiency (E) is defined as the ratio of the sorbent utilization with a given sorbent ( $U_i$ ) to that of hydrated lime A ( $U_a$ ) at the same conditions, calculated as follows:

$$E = 100 \times (U_i/U_a) \quad (24)$$

$U_i$  = Utilization of sorbent i, %

$U_a$  = Utilization of sorbent a, %

where,

$U_i$  or  $U_a$  is defined as,

$$U = \frac{\Delta SO_2 \times 100}{(L/74.08 + 0.5N/40.08)} \quad (25)$$

where,

$\Delta SO_2$  =  $SO_2$  capture, lb mol/hr

L = sorbent feed, lb/hr

N = NaOH feed, lb/hr

The data in the figure are averages of multiple observations of the relative efficiency, i.e., from the sorbent utilizations across the humidifier and system (humidifier + baghouse) at each of the three test conditions. The test conditions were:

Condition 1 - Ca/S = 1:1 mol/mol, NaOH/sorbent 0.0 mol ratio, 25°F approach.

Condition 2 - Ca/S = 1:1 mol/mol, NaOH/sorbent 0.2 mol ratio, 25°F approach.

Condition 3 - Ca/S = 2:1 mol/mol, NaOH/sorbent 0.2 mol ratio, 25°F approach.

The calculation of relative sorbent efficiency is detailed in Reference 5. In this test program, the NaOH additive ratio was controlled on a weight basis (1b NaOH/1b total sorbent), since sorbents of varying calcium contents were tested.

Figure 105 shows that sorbent selection is important for the Coolside process. The worst performing sorbent provided 35% (relative) lower sorbent utilization on average than the best sorbent (Lime A). Calcitic hydrated limes were generally superior to dolomitic hydrates. For the comparison shown, the different hydrated dolomitic and calcitic limes are compared based on their SO<sub>2</sub> removal performance per unit sorbent weight.

The variation in sorbent physical properties partially explains observed differences in sorbent performance. For hydrated calcitic limes, SO<sub>2</sub> capture increased somewhat with increasing BET surface area (Figure 106) and porosity, but the impact of these variables was reduced with NaOH additive injection. The surface area variation for these limes was from 14 to 22 m<sup>2</sup>/g, as measured by N<sub>2</sub> adsorption on the single-point BET method. The porosity varied from 19 to 30% for pores having diameters in the range of 1.5 to 300 nm, as measured by N<sub>2</sub> adsorption.

#### Once-Through Process Simulation Tests

The once-through simulation tests were made with the two sorbents selected for use at Edgewater. Tests were conducted over a wide range of process conditions, including Edgewater site-specific conditions. NaOH was used as the additive in the humidification water, since this was to be used at Edgewater. The tests with hydrated lime B were based on a Box-Behnken statistical experimental design in five variables: approach to adiabatic saturation (25 to 45°F), Ca/S mol ratio (0.75 to 2.25), Na/Ca ratio (0 to 0.2), inlet SO<sub>2</sub> content (500 to 2500 ppm wet basis), and inlet flue gas temperature (270 to 330°F). The tests with hydrated lime A were made with Ca/S and Na/Ca ratios as variables at constant 25°F approach, 300°F inlet temperature and 1500 ppm (wet) SO<sub>2</sub> content.

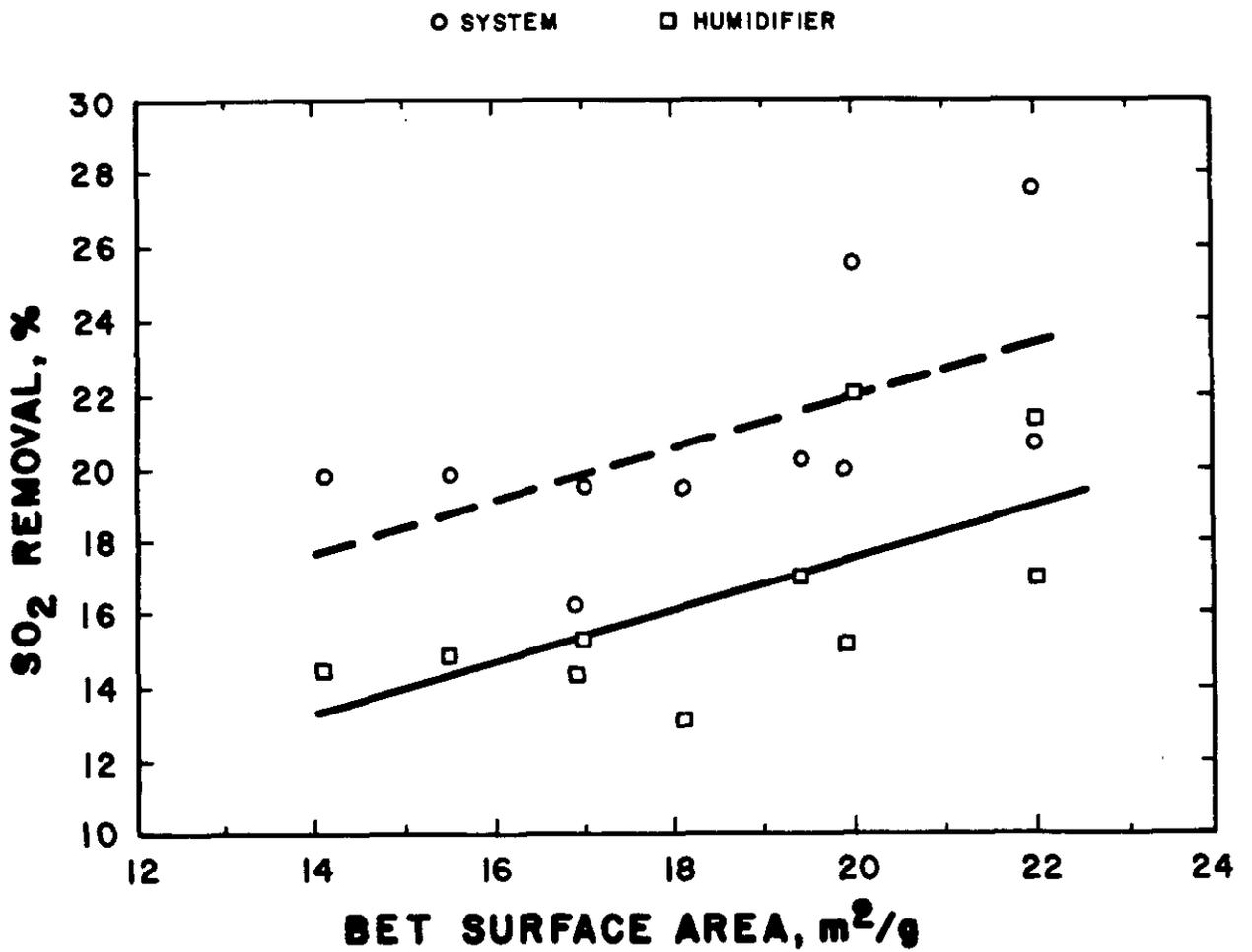


Figure 106. Effect of BET surface area of hydrated calcitic limes in Coolside process tests at Ca/S = 1, Na/Ca = 0, 25°F approach.

The test results showed that process  $\text{SO}_2$  removal increases strongly with closer approach to saturation and with increasing Ca/S and Na/Ca ratios, as observed previously.<sup>3</sup> Flue gas inlet temperature and  $\text{SO}_2$  content had secondary effects on  $\text{SO}_2$  removal. Based on the pilot data from the once-through simulation tests, statistical correlations were developed for  $\text{SO}_2$  removal as a function of key process variables. Correlation equations are given in Reference 1.

The statistical correlations indicated that the process must be operated at the closest practical approach to adiabatic saturation to ensure a maximum level of  $\text{SO}_2$  removal at given Ca/S and Na/Ca ratios. At 2.25 Ca/S mol ratio with no additive, the predicted system  $\text{SO}_2$  removal roughly doubles as the approach is changed from 45 to 25°F. Based on this, the target approach to saturation in the Edgewater demonstration was 20 to 25°F.

It was expected that Coolside process performance at Edgewater would differ from the predictions of the statistical correlations of pilot data. This is because the Edgewater Station uses an ESP for particulate removal, in contrast to the baghouse used in the pilot plant. Further, there are differences between the Edgewater and pilot plant humidifier design (e.g., flue gas velocity, sorbent particle distribution in the flue gas, atomizer design, droplet size distribution and drying time, etc.). However, the statistical correlations still provide information on sensitivity to key process variables as well as basis for correlating demonstration data with the extensive pilot plant data base.

Figure 107 shows a contour plot of system  $\text{SO}_2$  removal (humidifier + baghouse) as a function of Ca/S and Na/Ca mol ratios, as predicted by the correlation for hydrated lime G. The common conditions were 25°F approach to saturation, 1620 ppm  $\text{SO}_2$  (dry), and 300°F flue gas inlet temperature. The correlation predicts that the attainable  $\text{SO}_2$  removal levels without NaOH injection would be less than 50% at Ca/S mol ratios up to 2.25. NaOH injection of up to 0.2 Na/Ca ratio would expand the attainable range of  $\text{SO}_2$  removals and reduce sorbent requirements, indicating the importance of sodium additive injection for high  $\text{SO}_2$  removals with this lime.

Figure 108 shows a contour plot of system  $\text{SO}_2$  removal as a function of Ca/S and Na/Ca ratios, as predicted by the correlation for hydrated lime A. The plot

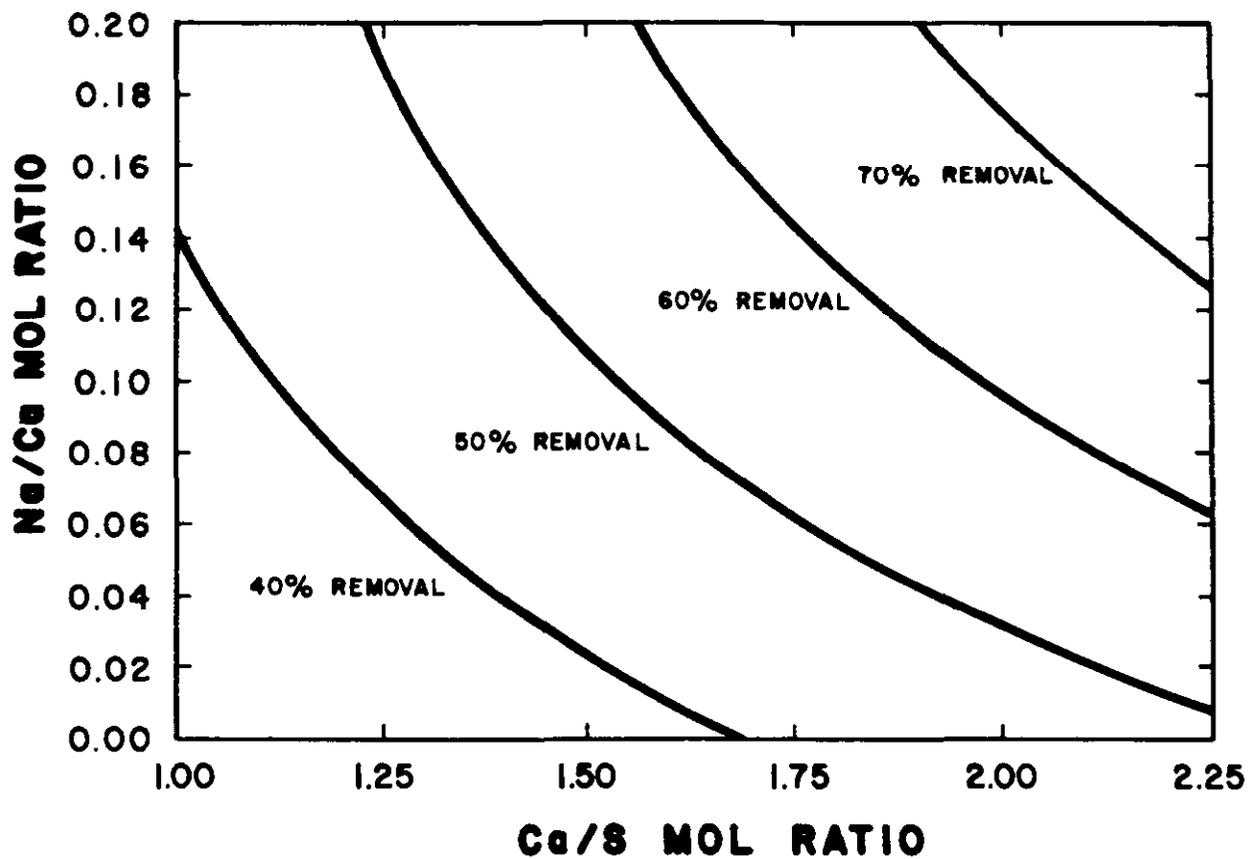


Figure 107. System SO<sub>2</sub> removal as a function of Ca/S and Na/Ca based statistical correlation of pilot-scale data with hydrated lime G for 25°F approach to saturation, 300°F inlet temperature, and 1620 ppm SO<sub>2</sub> concentration.

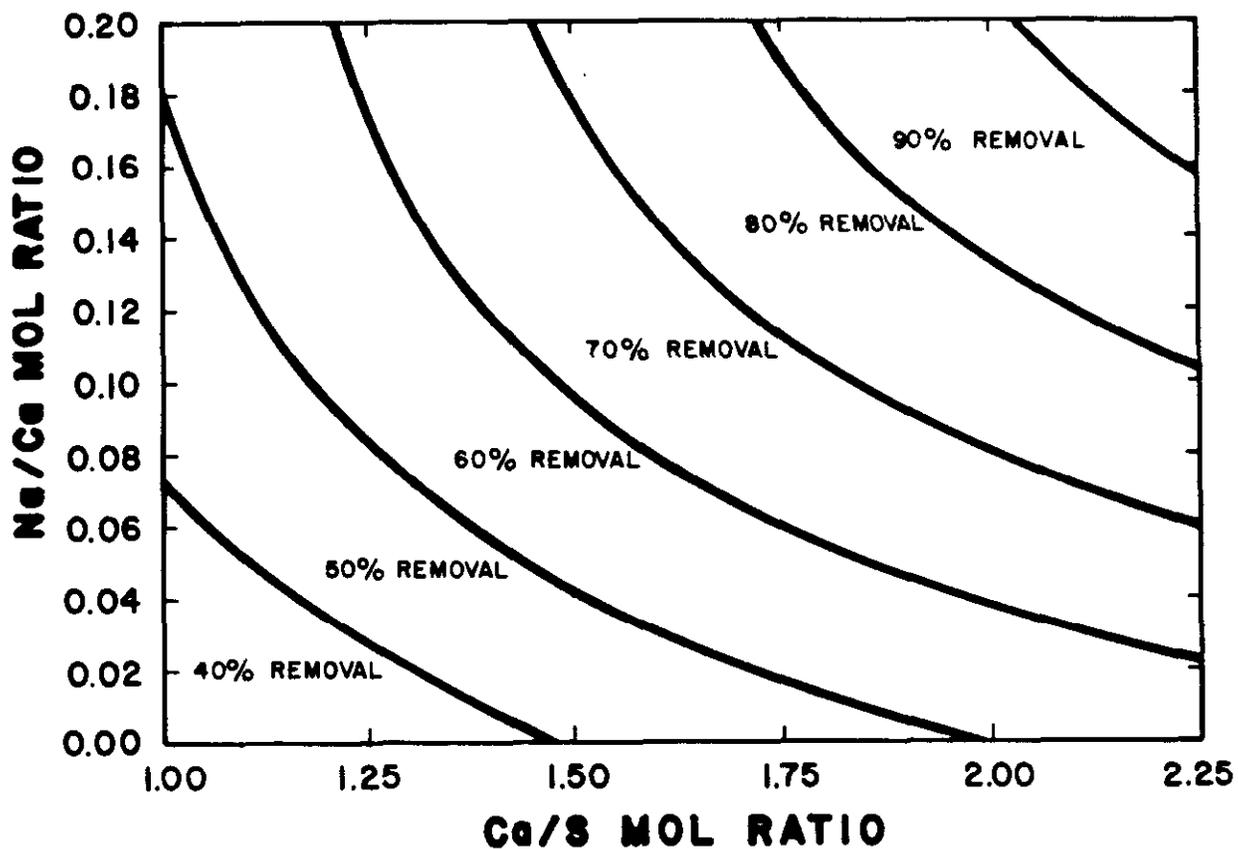


Figure 108. System SO<sub>2</sub> removal as a function of Ca/S and Na/Ca based on statistical correlation of pilot-scale data with hydrated lime A for 25°F approach to saturation, 300°F inlet temperature, and 1620 ppm SO<sub>2</sub> concentration.

shows that with this more reactive sorbent, significantly less sorbent and additive are required to attain the same target SO<sub>2</sub> removals. For example, in order to achieve a 60% SO<sub>2</sub> reduction with 0.1 Na/Ca ratio, the predicted sorbent and additive requirements are about 20% lower with hydrated lime A than with hydrated lime G.

The correlation for hydrated lime G predicts the effects of flue gas inlet temperature and SO<sub>2</sub> content on process SO<sub>2</sub> removal. SO<sub>2</sub> removal increases moderately with increasing flue gas temperature. For example, the predicted system removal increases by about 4% (absolute) with a 50°F increase in inlet temperature. This effect may be due to the increased quantity of water spray required to humidify the hotter flue gas, which results in more droplet-sorbent interactions. Previously published work<sup>26</sup> showed that these interactions enhance SO<sub>2</sub> removal in the humidifier. SO<sub>2</sub> removal decreases with increasing flue gas SO<sub>2</sub> content. For example, predicted system removal decreases by about 4% (absolute) with a 1000 ppm increase in SO<sub>2</sub> content. This effect is due to the decreasing ratio of water droplets to sorbent particles with increasing SO<sub>2</sub> content at the same Ca/S ratio.

### Recycle Process Simulation Tests

Pilot tests made under Edgewater site-specific conditions demonstrated a significant positive effect of sorbent recycle on sorbent utilization. These tests involved simultaneous injection of hydrated lime and recycle sorbent as shown in the process schematic (Figure 104). Based on the test results with the hydrated lime G, fresh sorbent and additive usage could be reduced significantly by sorbent recycle, although the attainable reduction depends on the capacity of the particulate collector, normally an ESP for power stations, and the subsequent solids handling equipment.

In all the recycle tests, hydrated lime G was fed with the fresh Ca/S mol ratio fixed at 1.0. The recycle tests were made with and without additive (NaOH) injection. For the tests with additive injection, the NaOH injection rate was fixed at 0.2 Na/Ca mol ratio based on the fresh hydrated lime feed. The approach to saturation was 25°F. Recycle sorbent was injected simultaneously with the fresh hydrated lime at recycle ratios (R) of 0.5, 1.0 and 1.8, where R is defined

as (lb recycle sorbent)/(lb fresh hydrated lime and fly ash). Steady-state continuous recycle was simulated in the pilot tests by making successive sets of runs with batchwise recycle at the same recycle ratio. The test method is described in detail in Reference 25.

Figure 104 shows observed SO<sub>2</sub> removal in the pilot tests made with NaOH injection at 0.2 Na/Ca as a function of recycle ratio (R). Sorbent recycle significantly increased measured system (humidifier + baghouse) SO<sub>2</sub> removals, from 41% with no recycle to over 60% with 1.8 recycle ratio. The results indicate that the recycle sorbent has substantial reactivity and capacity for additional SO<sub>2</sub> capture.

The recycle simulation results indicate that the recycle sorbent was almost as active as fresh sorbent when compared on the basis of available alkali. Figure 110 shows SO<sub>2</sub> removals with recycle were somewhat lower than for once-through operation at comparable available alkali/S mol ratios (A/S), defined as

$$A/S = (F_c + F_N/2 + R_c + R_N/2 - R_s)/S \quad (26)$$

where,

F<sub>c</sub> = mol/hr calcium in feed lime

F<sub>N</sub> = mol/hr sodium in humidification water

R<sub>c</sub> = mol/hr calcium in recycle solids

R<sub>N</sub> = mol/hr sodium in recycle solids

R<sub>s</sub> = mol/hr sulfur in the recycle solids

S = mol/hr SO<sub>2</sub> in the flue gas

The once-through data in the figure were calculated using the once-through correlation derived for hydrated lime G. The somewhat lower activity of the recycle sorbent may be partly due to the fact that some of the available calcium in the recycle material had been converted to CaCO<sub>3</sub>, an inactive Coolside sorbent. For example, for the recycle sorbent feed to the first set of recycle

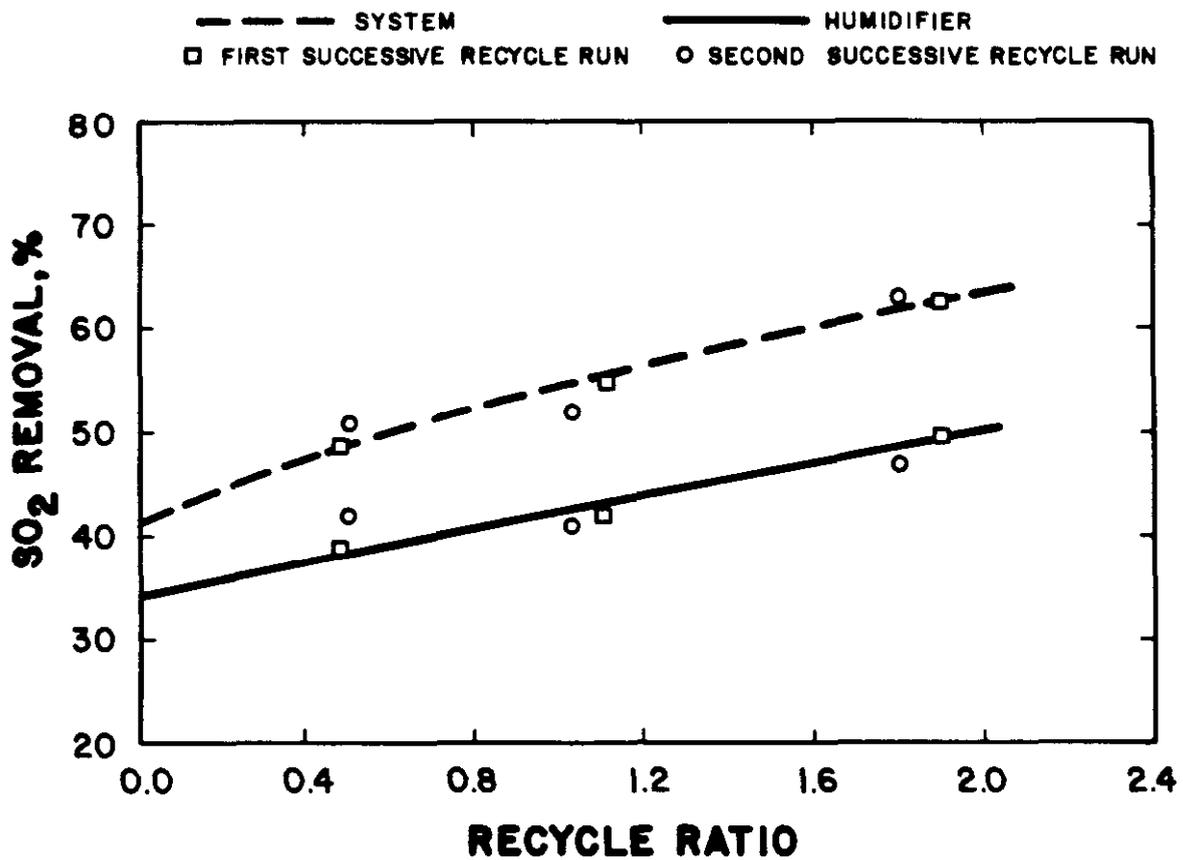


Figure 109. SO<sub>2</sub> removal as a function of recycle ratio at 24-26°F approach to saturation, 0.92-1.04 fresh Ca/S mol ratio and 0.19-0.21 fresh Na/fresh Ca mol ratio.

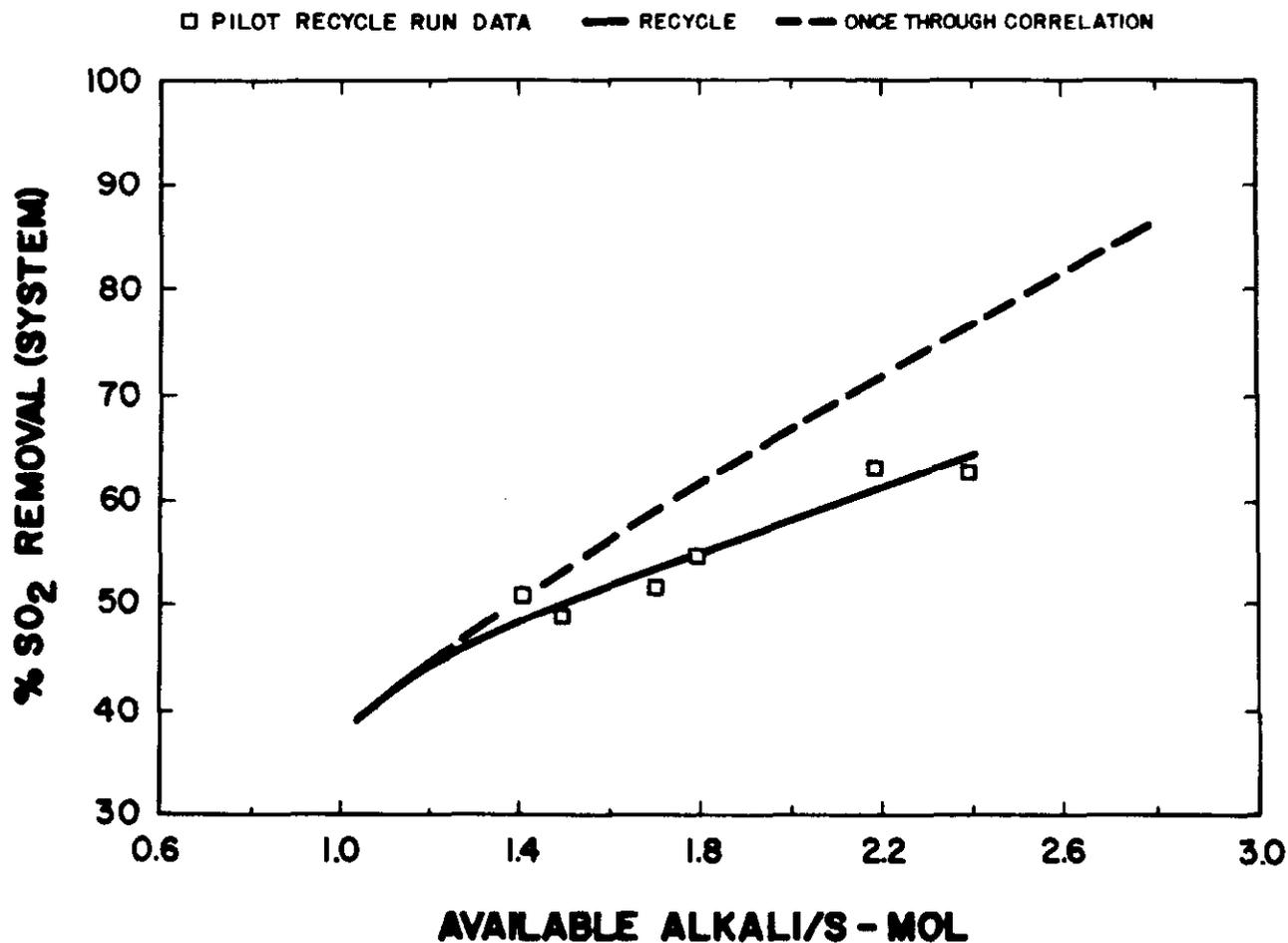


Figure 110. Comparison of system SO<sub>2</sub> removals with recycle and once-through operation as a function of available alkali with hydrated lime G as the fresh sorbent, at 25°F approach, 300°F inlet temperature, 2000 ppm SO<sub>2</sub>, and 0.2/1 fresh Na/fresh Ca mol ratio. (Recycle runs at constant 1/1 fresh Ca/S mol ratio.)

runs, about 10% of the total available calcium in the recycle sorbent (i.e., all Ca not reacted with sulfur) was in the form of  $\text{CaCO}_3$ .

In the successive pilot runs with batchwise recycle, the conditions for steady-state recycle were approached but not quite attained. For steady-state continuous recycle, sorbent utilization would be somewhat higher and  $\text{SO}_2$  removal somewhat lower than those measured. The table below gives estimates of steady-state system  $\text{SO}_2$  removal and sorbent utilization for 1.0 fresh Ca/S ratio with Black River hydrated lime, 0.2 fresh Na/Ca mol ratio, and 25°F approach to saturation, based on pilot data and material balance calculations.

<u>Recycle Ratio</u>	<u><math>\text{SO}_2</math> Removal (System), %</u>	<u>Overall Sorbent Utilization (System), %</u>
0	41	37
0.5	48	43
1	50	46
1.8	54	49

These results are based on the assumption that the single-pass conversion of available sorbent and additive (Ca and Na not tied up with sulfur) is constant at each recycle ratio and equal to the average value observed in the pilot runs. The method of projecting steady state conditions is detailed in Reference 25.

The above table shows that sorbent recycle can substantially reduce sorbent and additive requirements. In order to achieve the same 54%  $\text{SO}_2$  removal without recycle, a Ca/S mol ratio of 1.4 and a Na/Ca mol ratio of 0.2 would be required, based on the once-through correlation for hydrated lime G. This indicates that a sorbent recycle ratio of 1.8 has the potential to reduce the fresh sorbent and additive requirements by about 30%. The level of recycle possible at a given station will depend on capacity and operability of the particulate collector and particulate handling systems.

SECTION 7  
EVALUATION OF COOLSIDE SOLID WASTE PROPERTIES

Consol's activities concerning the evaluation of Coolside solid waste properties can be divided into three areas:

1. Prior to the availability of waste product from the Edgewater Station Coolside demonstration program, Consol Coolside pilot plant wastes were characterized and evaluated to address handling, transportation, and disposal concerns. The pilot plant was operated at Edgewater design conditions to produce the wastes evaluated. This work was performed in support of plans to dispose of the waste that would be produced in the demonstration test.
2. Consol issued a subcontract to Baker/TSA, Inc. to recommend landfilling procedures for Coolside waste. Baker/TSA's recommendations were based upon Consol's characterization of the Coolside pilot plant waste.
3. Consol characterized and evaluated Coolside waste produced during the Edgewater demonstration program. The properties examined were relevant to handling, transportation and disposal. In addition, this material was evaluated in regard to two potential by-product uses.

The first two activities listed above were supported in part by the DOE Cooperative Agreement. The third activity was funded solely by Consol. The results of the third activity are included here because they provide a valuable addition to the results of the first two activities. Summaries of these three activities appear below under separate subheadings. The complete results of the first two activities were reported previously.<sup>28,29,30,31</sup> The complete results of the third activity<sup>32</sup> appear in Appendix F.

## COOLSIDE WASTE FROM THE CONSOL PILOT PLANT

The purpose of this work was to determine the properties of the Coolside solid waste in order to support plans to landfill the waste that would be produced during the demonstration test. The solid-waste samples used in this work originally were produced in the Consol Coolside pilot plant at mol ratio of 2 Ca/S and 25°F approach to adiabatic saturation. Samples were examined that were produced both with NaOH/Ca(OH)<sub>2</sub> mol ratio of 0.19 and without the sodium additive.

The waste characterization program consisted of seven elements. The first six were laboratory tests and the seventh was a small-scale field test in which the waste was exposed to the weather for about six months.

### Optimum Moisture

The optimum moisture content to achieve the maximum dry bulk density of the Coolside waste (produced with the sodium hydroxide additive) is about 30 wt % dry basis, i.e., 30 lbs water to 100 lbs dry solids. At this moisture content, the waste has the appearance of moist sand. At the optimum moisture content, the dry, compacted bulk density (ASTM D-698) is 66.8 lb/ft<sup>3</sup>. The loose and tapped bulk densities of the waste containing the optimum moisture are 30.5 lb/ft<sup>3</sup> and 39.4 lb/ft<sup>3</sup>, respectively.

### Compressive Strength

Unconfined compressive strengths were measured on Coolside wastes produced both with and without the sodium hydroxide additive. Compressive strengths increase with increasing moisture content (over the range 20 to 32 weight percent, dry basis) and curing time (up to 28 days). After 28 days curing at the optimum moisture content, measurements indicated that the compressive strength (251 psi) of the wetted waste is adequate to permit landfill disposal. In the absence of the sodium hydroxide additive, the compressive strength at the same conditions was 200 psi.

## Toxicity

Leachate toxicity was measured using the U.S. Environmental Protection Agency (EPA) Extraction Procedure method<sup>33</sup> (EP procedure) and by the same procedure using only deionized water instead of aqueous acetic acid. Wastes produced both with and without the sodium hydroxide additive were tested.<sup>32</sup> For both leaching methods, both with and without the sodium hydroxide additive, the leachates are within Resource Conservation and Recovery Act (RCRA) limits and the concentrations of trace elements, sulfate and total dissolved solids (TDS) in the leachates are less than thirty times the EPA primary and secondary drinking water standards (Ohio EPA requirements). Good material balances (based on measured and calculated TDS) and charge balances (based on concentrations of anions and cations) were obtained in these leachate composition measurements. Nitrate, fluoride and chloride could not be determined on the standard EP leachates because of interferences from the acetate ion. However, their concentrations were well below Ohio standards when the deionized water leaching method was used. The leachates from the waste produced without the sodium hydroxide additive have reduced sodium, sulfate and total dissolved solids concentrations, but otherwise are similar to those from the samples produced with the additive. Results from a Coolside waste leachate test, on a waste sample which was first cured for 28 days and ground to minus 200 mesh (minus 75  $\mu\text{m}$ ) prior to leaching, indicate that curing did not significantly affect EPA leachability results. It is expected that fully consolidated, cured waste would have a permeability in the  $10^{-5}$  to  $10^{-6}$  cm/sec range (see below). The low permeability is expected to reduce the leachability of the waste.

## Permeability

Permeability coefficients ranging from  $10^{-5}$  to  $10^{-6}$  cm/sec were obtained using ASTM D-698 to prepare Coolside waste (produced with the sodium hydroxide additive) containing the optimum moisture. Materials with permeability coefficients in this range generally are considered to be suitable for landfill disposal. Curing time appears to reduce permeability slightly.

### Wetting Temperature Rise

Only a slight temperature increase ( $\leq 10^{\circ}\text{F}$ ) was observed in Coolside waste after mixing with 10 to 30 weight percent added water. Thus, wetting Coolside waste is not expected to result in handling problems caused by a temperature rise as can occur with LIMB waste. The reason for the low temperature rise is that Coolside process wastes do not contain unhydrated  $\text{CaO}$ , whereas LIMB wastes do.

### Composition

Three techniques (thermogravimetric analysis, Fourier-transform infrared spectroscopy, and lime index measurement) were used to characterize changes in the chemical component composition of the waste before and after curing. The concentration of  $\text{Ca}(\text{OH})_2$  was observed to decrease with increasing curing time (and thus with increasing compressive strength), indicating that pozzolanic reactions proceeded during curing. One cured sample was analyzed by X-ray diffraction (XRD) and found to contain ettringite ( $\text{Ca}_6\text{Al}_2(\text{SO}_4, \text{SiO}_4, \text{CO}_3)_3(\text{OH})_{12} \cdot 28\text{H}_2\text{O}$ ). Though the only representative Coolside waste sample analyzed by XRD was not produced at simulated Edgewater flue gas conditions, the presence of ettringite indicates that pozzolanic reactions do indeed occur as Coolside waste cures.

### Weathering Tests

Two waste piles were exposed to the weather for six months. Both piles consisted of Coolside waste with sodium additive. The piles were made by manually compacting Coolside wastes with moisture contents of 20 and 33.3%. The piles remained soft, loose, and permeable for the duration of the six-month test. The high permeabilities compared to a standard landfill of the piles would tend to reduce the contact time of the percolation water with the solid, but at the same time would tend to increase the ratio of percolation water to run-off water. The experimental design was such that the drainage water quality results obtained in this study are probably worst case values. In all cases, the concentrations of the trace elements in the drainage water were less than 30 times the U.S. EPA primary and secondary drinking water standards. Though several of the early drainage samples had high TDS and sulfate concentrations, all later drainage

samples had acceptable levels. The pH of the drainage water varied widely (7.4 to 12.4) depending on the relative amount of percolation and run-off and on atmospheric CO<sub>2</sub> absorption. Sodium concentrations in the bulk waste piles were reduced by about 90% by the end of the approximately six-month tests, confirming that much of the precipitation percolated through the unconsolidated shallow waste piles.

These results suggest that Coolside waste should be suitable for landfill disposal. The complete results and details of the work described above were issued as part of contract reporting in a document, "Coolside Waste Management Studies-- Final Report", that was issued September 1988.<sup>32</sup> Some of these results also were reported externally.<sup>33,34</sup>

#### LANDFILL PROCEDURE RECOMMENDATIONS

Consol issued a subcontract to Baker/TSA, Inc. to conduct a regulatory review, to evaluate the properties of Coolside waste as reported by Consol,<sup>32</sup> and to recommend landfilling procedures. Baker concluded<sup>35</sup> that Coolside waste should be handled as a solid waste rather than a hazardous waste. Baker also provided general guidelines for the evaluation of landfills and for specific landfill practices. Baker included in their report grain-size analyses of Coolside waste performed by Consol. The final report by Baker<sup>35</sup> was included by Consol in contract reporting.

#### COOLSIDE WASTE FROM THE EDGEWATER STATION

The waste materials discussed above were produced in the Consol Coolside pilot plant. Consol also evaluated Coolside waste produced during the Edgewater demonstration program.<sup>36</sup> Though this work was funded solely by Consol and was not part of the DOE Cooperative Agreement, the results are summarized here because they provide a valuable addition to prior studies. Complete results appear in Appendix F. The Edgewater waste material was produced at the following conditions: Ca/S mol ratio of 1.4, NaOH/Ca(OH)<sub>2</sub> mol ratio of 0.20, and 20°F approach to adiabatic saturation. Note that the pilot plant wastes, discussed above, were produced at a Ca/S molar ratio of 2.0.

In many respects, the Edgewater waste has superior structural properties when compared with the pilot plant wastes. For example, though optimum moisture content (33 versus 30 weight percent, dry basis) and particle size ( $D_{50} = 7.6$  versus  $5.5 \mu\text{m}$ ) are similar, the Edgewater waste has a higher maximum dry bulk density (81.8 versus 66.8 lb/ft<sup>3</sup>), a higher unconfined compressive strength (655 versus 251 psi), and a lower permeability coefficient ( $1.7 \times 10^{-8}$  to  $2.0 \times 10^{-7}$  cm/sec) than the pilot plant waste. The above strength and permeability values were obtained on specimens cured for 28 days with the optimum moisture content.

Leaching tests indicated that the Edgewater wastes can be classified as non-hazardous.

Edgewater Coolside waste also was evaluated for use in acid mine drainage (AMD) treatment. This material appears to be a suitable replacement for conventional hydrated lime in AMD treatment in terms of iron oxidation and neutralization rates, sludge settling properties and effluent water quality.

Edgewater Coolside waste was pelletized and the products were characterized in terms of strength, density, particle size and leachability. Pelletization enhances handleability for transportation and reduces waste leachability. The pelletized products, after curing, have potential to be used as synthetic aggregates for road base construction.

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**DATA AND SAMPLE ANALYSES**

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## LIME SURFACE AREA AND TGA DATA

Hydrated Lime A

Sample Date	Sample Number	Surface Area m <sup>2</sup> /g	TGA Data					
			wt 1	wt 2	wt 1	wt 2	wt % Ca(OH) <sub>2</sub>	wt % CaCO <sub>3</sub>
08/29/89	082989-LM103	23.05	98.969	76.569	76.569	75.336	93.1	2.83
09/08/89	090889-LM106	23.99	98.996	76.333	76.333	75.649	94.2	1.57
09/08/89	090889-LM108	23.59	99.516	77.061	77.061	75.248	92.8	4.14
09/11/89	091189-LM110	23.18	98.977	76.348	76.348	75.446	94.0	2.07
09/12/89	091289-LM112	23.13	98.967	76.154	76.154	75.419	94.8	1.69
09/25/89	092589-LM119	23.56	99.009	77.044	77.044	75.426	91.2	3.72
09/26/89	092689-LM121	23.60	99.239	76.833	76.833	75.713	92.9	2.57
09/28/89	092889-LM124	22.75	99.335	77.177	77.177	75.923	91.7	2.87
10/03/89	100389-LM126	22.63	99.141	76.495	76.495	75.589	93.9	2.08
10/04/89	100489-LM128	23.53	99.255	77.250	77.250	75.849	91.2	3.21
10/05/89	100589-LM130	23.20	99.161	76.577	76.577	75.802	93.7	1.78
10/10/89	101089-LM132	23.37	99.345	76.717	76.717	75.898	93.7	1.87
10/11/89	101189-LM134	23.16	99.132	76.761	76.761	76.054	92.8	1.62
10/12/89	101289-LM136	23.07	98.914	76.635			92.6	
10/13/89	101389-LM138	21.87	99.006	76.702			92.7	
10/18/89	101889-LM140	23.61	98.942	76.688			92.5	
10/19/89	101989-LM142	22.13	98.998	77.134	76.908	75.024	90.8	4.33
10/26/89	102689-LM144	22.24	99.322	77.320	77.299	75.698	91.1	3.67
11/01/89	110189-LM146	23.97	99.184	76.630	76.630	75.647	93.5	2.25
11/02/89	110289-LM148	24.12	99.178	76.534	76.534	75.857	93.9	1.55
11/13/89	111389-LM150	23.77	99.368	76.748	76.748	75.967	93.6	1.79
11/28/89	112889-LM152	22.45	99.156	76.724	76.729	75.635	93.0	2.51
11/30/89	113089-LM154	23.36	99.319	76.664	76.664	75.922	93.8	1.70
12/04/89	120489-C-156	23.12	99.333	76.691	76.691	75.508	93.8	2.71
12/05/89	120589-LM158	23.59	99.228	76.762	76.762	75.812	93.1	2.18
12/11/89	121189-LM162	23.51	99.282	76.736	76.736	75.847	93.4	2.04
01/11/90	011190	22.77	99.349	77.097	77.097	75.878	92.1	2.79
AVERAGES		23.20					92.97	2.48
STD. DEV.		0.563					1.017	0.810

### LIME SURFACE AREA AND TGA DATA

Hydrated Lime G

Sample Date	Sample Number	Surface Area m <sup>2</sup> /g	TGA Data					
			wt 1	wt 2	wt 1	wt 2	wt % Ca(OH) <sub>2</sub>	wt % CaCO <sub>3</sub>
01/19/90	011990-1515	17.53	99.023	78.388	78.388	76.794	85.7	3.66
01/20/90	012090-0225	16.96	98.893	77.775	77.780	76.619	87.8	2.67
01/20/90	012090-1035	17.23	98.594	77.379	77.379	76.173	88.5	2.78
01/20/90	012090-1915	16.99	98.789	77.591	77.591	76.519	88.3	2.47
01/21/90	012190-0035	16.83	98.845	77.631	77.631	76.817	88.3	1.87
01/21/90	012190-0650	17.14	98.890	77.734	77.734	76.763	88.0	2.23
01/22/90	012290-0605	16.89	98.978	77.838	77.838	76.727	87.8	2.55
01/23/90	012390	16.93	98.832	77.456	77.456	76.365	89.0	2.51
01/23/90	012390-0200	15.62	98.824	77.511	77.511	76.345	88.7	2.68
01/24/90	012490	16.72	98.909	77.728	77.719	76.543	88.1	2.70
01/24/90	012490-0415	16.42	98.827	77.441	77.441	76.509	89.0	2.14
01/25/90	012590	16.19	98.934	77.558	77.558	76.668	88.9	2.05
01/25/90	012590-0215	17.21	98.870	77.706	77.706	76.566	88.0	2.62
01/25/90	012590-0450	17.04	98.884	77.478	77.478	76.568	89.0	2.09
01/25/90	012590-2345	16.99	98.890	77.501	77.501	76.493	89.0	2.32
01/26/90	012690-0630	17.12	98.881	77.612	77.612	76.422	88.5	2.74
01/27/90	012790-0325	16.94	98.833	77.750	77.750	76.610	87.7	2.62
01/27/90	012790-0700	17.50	99.031	77.924	77.924	76.897	87.7	2.36
02/01/90	020190-0100	17.00	98.861	77.675	77.675	76.799	88.1	2.02
02/02/90	020290-0615	17.18	98.822	77.635	77.635	76.634	88.2	2.30
02/02/90	020290-1340	14.71	98.727	77.708	77.739	76.451	87.6	2.97
02/02/90	020290-1940	16.25	98.846	77.602	77.602	76.722	88.4	2.02
02/03/90	020390-0230	17.04	98.875	77.555	77.555	76.614	88.7	2.16
02/03/90	020390-0615	17.34	98.854	77.580	77.580	76.606	88.5	2.24
02/03/90	020390-1457	16.50	98.827	77.704	77.704	76.667	87.9	2.39
02/03/90	020390-1820	17.20	98.907	77.673	77.673	76.866	88.3	1.86
02/06/90	020690-0020	17.22	98.889	77.600	77.600	76.705	88.5	2.06
02/06/90	020690-0545	17.09	98.808	77.473	77.473	76.431	88.8	2.40
02/07/90	020790-0100	17.01	98.989	78.222	78.222	77.394	86.3	1.90
02/07/90	020790-0635	16.48	98.959	78.285	78.285	77.346	85.9	2.16
02/08/90	020890-0615	16.13	98.786	77.455	77.455	76.585	88.8	2.00
02/08/90	020890-1450	16.51	98.600	77.105	77.105	75.694	89.7	3.25
02/09/90	020990-0100	16.71	98.829	77.962	77.962	77.047	86.8	2.11
02/09/90	020990-0615	16.49	98.884	78.501	78.501	76.978	84.8	3.50
02/10/90	021090-0100	17.16	98.825	77.849	77.849	76.620	87.3	2.83
02/10/90	021090-0620	16.66	98.875	78.020	78.020	77.025	86.8	2.29
02/10/90	021090-1025	16.04	98.879	78.879	78.879	77.596	83.2	2.95
02/10/90	021090-2305	17.54	98.833	77.299	77.299	76.203	89.6	2.52
02/11/90	021190-0355	16.64	98.991	77.509	77.509	76.334	89.3	2.70
02/12/90	021290-0030	15.58	98.720	78.654	78.654	77.013	83.6	3.78
02/12/90	021290-0550	17.60	98.924	77.480	77.480	76.490	89.2	2.28
02/13/90	021390-0055	15.58	98.836	77.180	77.180	76.042	90.1	2.62
02/14/90	021490-0300	15.40	98.736	77.232	77.232	76.216	89.6	2.34
<b>AVERAGES</b>		16.73					87.95	2.48
<b>STD. DEV.</b>		0.626					1.469	0.451

## MALVERN PARTICLE SIZE ANALYSIS

SAMPLE DATE	10/13/89	12/05/89	01/25/90	01/27/90	02/09/90
HYDRATED LIME	A	A	G	G	G
ANALYSIS NUMBER	900089	900090	900695	900696	900930
SIZE RANGE	PERCENT IN SIZE RANGE				
Under 1.2 $\mu\text{m}$	7.3	2.6	0.6	1.0	0.5
1.2 to 1.4 $\mu\text{m}$	0.7	0.6	0.2	0.2	0.2
1.4 to 1.6 $\mu\text{m}$	0.9	0.9	0.5	0.5	0.4
1.6 to 1.9 $\mu\text{m}$	2.0	1.2	0.8	0.6	0.5
1.9 to 2.2 $\mu\text{m}$	3.6	2.0	1.7	1.2	1.2
2.2 to 2.6 $\mu\text{m}$	6.6	3.6	3.9	3.8	2.7
2.6 to 3.0 $\mu\text{m}$	9.0	4.8	5.6	6.4	4.1
3.0 to 3.4 $\mu\text{m}$	9.1	5.8	6.8	7.3	4.9
3.4 to 4.0 $\mu\text{m}$	8.0	6.4	7.7	7.6	5.5
4.0 to 4.6 $\mu\text{m}$	6.8	6.8	8.5	7.9	6.3
4.6 to 5.3 $\mu\text{m}$	6.2	7.7	9.6	9.4	7.9
5.3 to 6.2 $\mu\text{m}$	5.9	8.5	10.7	11.1	9.8
6.2 to 7.2 $\mu\text{m}$	5.1	8.5	10.6	11.4	10.8
7.2 to 8.3 $\mu\text{m}$	4.3	7.9	9.9	10.2	10.7
8.3 to 9.6 $\mu\text{m}$	3.5	6.9	8.3	8.0	9.5
9.6 to 11.1 $\mu\text{m}$	2.8	5.7	6.0	5.5	7.8
11.1 to 12.9 $\mu\text{m}$	2.3	5.1	4.1	3.9	6.5
12.9 to 15.0 $\mu\text{m}$	1.9	4.1	2.6	2.5	4.6
15.0 to 17.4 $\mu\text{m}$	2.0	3.3	1.3	1.2	3.0
17.4 to 20.1 $\mu\text{m}$	2.3	2.5	0.4	0.4	1.9
20.1 to 23.3 $\mu\text{m}$	2.3	1.9	0.4	0.3	1.0
23.3 to 27.0 $\mu\text{m}$	1.6	1.4	0.1	0.1	0.3
27.0 to 31.3 $\mu\text{m}$	1.2	0.9	0.0	0.0	0.3
31.3 to 36.3 $\mu\text{m}$	1.5	0.5	0.0	0.0	0.2
36.3 to 42.1 $\mu\text{m}$	1.7	0.4	0.0	0.0	0.1
42.1 to 48.8 $\mu\text{m}$	1.4	0.4	0.0	0.0	0.0
48.8 to 56.6 $\mu\text{m}$	0.5	0.2	0.0	0.0	0.0
56.6 to 65.6 $\mu\text{m}$	0.0	0.0	0.0	0.0	0.0
65.6 to 76.0 $\mu\text{m}$	0.0	0.0	0.0	0.0	0.0
76.0 to 88.1 $\mu\text{m}$	0.0	0.0	0.0	0.0	0.0
88.1 to 102.1 $\mu\text{m}$	0.0	0.0	0.0	0.0	0.0
102.1 to 118.4 $\mu\text{m}$	0.0	0.0	0.0	0.0	0.0
AVG DIAM, $\mu\text{m}$	4.3	6.1	5.7	5.7	6.8

NaOH SAMPLES				
ANALY. NUMBER	SAMPLE DATE	SAMPLE NUMBER	SAMPLE DESCRIPTION	SOLUTION WT% NAOH
894921	09/19/89	091989	EDGEWATER CONC. NAOH	47.2
895021	10/11/89	1011891525	HUMIDIF WATER SAMPLE	1.63

**PROCESS RUN DATA - HYDRATED LIME A**

Test Date	Test Time (hrs)	Dura-tion (hrs)	Coal S (%)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	FI Gas (gal/kh)	Water/ Humidifier Inlet			Gas Analyzer Data			Ca/S, mol		Na/Ca (mol)	% SO2 Rem-oval	By-pass %	Recycle Feed, kbb/hr	Appx to Adiab Satur °F
					Inlet	Out-let	TV			SO2 ppm	O2 %	SO2 (corr)	SO2 ppm	O2 %	SO2 (corr)	Sys	Hum					
10/2 19:40	1.2	2.10	72.3	288	150	151	0.48	3.85	1469	6.72	2414	1088	7.89	1747	0.92	0.84	0.12	27.6	0	0	(31.8)	
10/2 22:00	1.3	2.10	44.0	277	145	148	0.75	3.57	1325	8.22	2468	899	9.52	1651	0.87	0.77	0.10	33.1	0	0	(26.6)	
10/3 01:50	4.2	2.18	43.8	273	145	147	0.78	3.56	1347	8.04	2470	957	9.29	1723	0.83	0.76	0.09	29.8	0	0	(25.6)	
10/3 13:16	1.3	2.18	70.2	278	145	145	0.50	3.88	1464	6.44	2355	726	7.55	1137	1.84	1.76	0.15	52.0	0	0	22.9	
10/4 13:37	1.0	2.39	61.1	275	143	145	0.56	3.77	1442	7.56	2538	786	8.97	1378	1.39	1.36	0.16	44.7	0	0	22.9	
10/4 16:07	2.2	2.39	44.9	270	143	145	0.76	3.65	1447	7.95	2632	804	9.34	1454	1.38	1.29	0.17	45.5	0	0	23.1	
10/4 22:17	1.7	2.39	42.7	272	143	145	0.77	3.66	1499	8.10	2765	867	9.40	1576	1.49	1.33	0.17	42.8	0	0	25.4	
10/5 00:00	6.2	2.45	42.9	271	142	144	0.76	3.70	1508	8.41	2767	858	9.34	1552	1.46	1.34	0.17	43.7	0	0	22.1	
10/5 00:58	2.3	2.45	67.5	288	142	141	0.41	4.36	1474	7.77	2641	658	9.27	1183	1.46	1.40	0.17	55.3	0	0	(18.6)	
10/5 13:58	1.2	2.45	61.7	289	142	142	0.41	4.30	1318	8.89	2611	618	10.1	1199	1.49	1.45	0.18	54.1	0	0	22.2	
10/5 15:58	2.7	2.45	62.7	291	142	142	0.40	4.35	1372	8.86	2713	641	10.1	1241	1.48	1.38	0.18	54.1	0	0	(22.1)	
10/5 19:18	2.2	2.45	72.7	290	142	141	0.38	4.41	1556	7.71	2774	751	8.97	1317	1.51	1.38	0.18	52.3	0	0	20.2	
10/6 10:39	3.7	2.62	67.4	292	142	141	0.42	4.45	1508	7.55	2652	691	8.60	1176	1.42	1.24	0.20	46.3	0	0	21.0	
10/6 16:52	3.5	2.62	64.2	290	142	146	0.69	4.35	1746	7.68	3103	936	9.17	1669	1.42	1.24	0.20	46.3	0	0	21.0	
10/7 00:02	3.5	2.77	46.1	276	142	145	0.84	3.76	1797	8.29	3369	1079	9.54	1985	1.31	1.12	0.20	41.4	0	0	(24.2)	
10/7 04:22	1.0	2.77	38.4	270	142	145	0.84	3.59	1684	8.36	3180	1004	9.65	1868	1.28	1.15	0.20	42.7	0	0	(24.0)	
10/7 07:12	1.7	2.77	38.0	265	142	142	0.80	3.59	1596	9.04	3210	926	10.1	1798	1.24	1.10	0.20	43.8	0	0	(22.6)	
10/10 18:25	1.5	2.69	64.8	281	155	155	0.67	3.50	1626	7.00	2733	1027	8.61	1746	1.41	1.43	0.19	36.1	0	0	31.4	
10/10 20:37	1.3	2.69	60.6	275	144	144	0.68	3.72	1640	7.10	2779	889	8.71	1524	1.43	1.43	0.18	44.7	0	0	20.3	
10/10 22:38	1.3	2.69	44.3	272	145	147	0.85	3.62	1503	8.53	2880	851	9.88	1615	1.43	1.38	0.21	44.2	0	0	(25.7)	
10/11 00:00	3.0	2.65	44.2	272	145	146	0.84	3.67	1491	8.39	2822	807	9.80	1520	1.44	1.40	0.20	46.0	0	0	25.9	
10/11 04:41	1.7	2.65	41.6	265	144	146	0.97	3.46	1459	8.44	2773	821	9.60	1518	1.44	1.42	0.19	44.4	0	0	25.0	
10/11 12:01	1.8	2.65	41.3	265	142	144	0.90	3.65	1408	8.96	2810	728	10.1	1413	1.44	1.40	0.21	49.5	0	0	(23.7)	
10/11 21:13	2.8	2.65	44.1	274	142	143	0.85	3.92	1567	7.67	2783	775	9.17	1382	1.46	1.43	0.20	50.0	0	0	21.4	
10/12 00:00	2.2	2.74	44.3	277	142	143	0.84	3.96	1560	7.60	2755	739	9.17	1318	1.41	1.45	0.20	52.4	0	0	(21.0)	
10/12 04:33	1.3	2.74	44.4	271	142	144	0.83	3.80	1487	8.11	2743	785	9.78	1475	1.41	1.46	0.20	45.4	0	0	22.7	
10/12 11:09	2.9	2.74	45.1	280	142	143	0.73	4.23	1514	7.91	2746	422	9.39	767	2.00	2.06	0.24	72.1	0	0	20.1	
10/12 19:46	1.1	2.74	64.6	292	147	148	0.76	4.29	1643	7.44	2863	488	8.83	845	1.95	1.93	0.21	70.5	0	0	19.7	
10/12 23:04	1.1	2.74	43.4	282	142	144	0.78	4.19	1574	7.75	2815	519	9.34	938	1.94	1.95	0.21	60.0	0	0	(23.7)	
10/13 00:00	1.1	2.66	43.4	280	142	143	0.79	4.16	1580	7.64	2800	511	9.25	918	2.01	1.97	0.21	65.5	0	0	(21.3)	
10/13 12:49	3.0	2.66	54.8	289	142	142	0.62	4.53	1637	6.96	2743	509	8.46	856	2.12	2.12	0.21	68.9	0	0	(20.5)	
10/16 17:25	1.3	2.63	42.1	284	144	146	0.76	4.23	1543	7.64	2794	448	9.22	802	2.07	2.06	0.22	71.2	0	0	22.4	
10/16 19:37	4.2	2.63	41.9	273	143	145	0.82	3.94	1524	7.68	2707	437	9.22	782	2.07	2.08	0.22	71.3	0	0	21.3	
10/17 00:01	6.2	2.60	41.3	269	143	145	0.86	3.88	1519	7.50	2659	420	9.10	744	2.08	2.11	0.21	71.6	0	0	21.8	
10/17 09:33	3.5	2.60	69.2	271	143	144	0.60	3.85	1581	6.71	2596	949	8.25	1569	1.08	1.12	0.21	40.1	0	0	22.6	

**PROCESS RUN DATA - HYDRATED LIME A**

Test Date	Test Duration (hrs)	Coal S (wt %)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	Water/Fl Gas (gal/lb)	Gas Analyzer Data						Na/Ca (mol)	% SO2 Rem-oval	By-pass %	Recycle Feed, klb/hr	Apply to Ashab Satur °F			
				Inlet	Out-	Humidifier			Humidifier Inlet		ESP Outlet		Ca/S, mol	Sys Hum								
				Inlet	Out-	Humidifier			SO2 ppm	O2 %	SO2 (corr)	SO2 ppm	O2 %	SO2 (corr)	Ca/S, mol	Sys Hum						
10/17 16:53	1.1	2.60	66.9	267	142	143	0.65	3.86	1618	5.85	2489	995	7.78	1584	1.08	1.17	0.12	0	(18.1)			
10/17 20:44	2.4	2.60	97.4	295	185	183	0.50	3.37	1693	6.17	2668	1398	7.78	2227	1.05	1.06	0.18	0	60.1			
10/18 11:35	1.3	2.90	93.4	284	152	152	0.42	3.92	1572	6.41	2524	618	7.70	979	1.73	2.05	0.24	0	29.3			
10/19 11:35	1.5	2.81	94.6	289	152	149	0.40	4.17	1622	6.01	2524	1019	7.65	1608	0.98	1.13	0.24	0	23.5			
10/19 13:25	1.1	2.81	94.9	295	155	156	0.39	4.15	1549	6.22	2450	1016	7.90	1634	0.97	1.15	0.24	0	30.3			
10/19 17:38	1.1	2.81	92.5	295	155	152	0.39	4.25	1751	5.56	2639	1091	7.54	1707	0.96	1.06	0.24	0	(26.5)			
10/26 10:26	2.2	1.79	97.6	307	144	189	0.46	4.62	1168	5.68	1774	787	7.51	1229	1.46	1.90	0.17	30.9	28	0	18.1	
10/30 18:25	1.7	1.69	86.2	297	150	180	0.54	4.07	1060	6.33	1689	928	7.93	1495	0.98	1.18	0.00	11.5	20	0	(25.4)	
10/30 22:16	1.8	1.69	46.5	275	145	147	0.83	3.63	1020	7.47	1781	681	8.79	1176	2.01	1.83	0.00	33.9	0	0	(24.4)	
10/31 02:07	3.7	1.60	46.2	269	145	147	0.87	3.53	923	7.59	1828	601	8.87	1044	2.19	2.13	0.00	36.4	0	0	22.6	
10/31 18:26	1.5	1.60	84.7	290	147	164	0.56	4.01	995	5.81	1525	691	7.43	1073	2.72	3.18	0.00	29.6	12	0	(22.1)	
11/1 00:00	2.1	1.53	44.7	265	140	142	0.81	3.63	823	7.84	1484	257	9.18	457	2.28	2.32	0.25	69.2	0	0	(19.9)	
11/1 02:41	1.3	1.53	45.2	264	140	142	0.82	3.61	832	7.73	1485	279	9.05	493	2.10	2.13	0.25	66.8	0	0	(19.5)	
11/1 14:20	2.2	1.53	43.3	258	142	144	0.98	3.41	806	7.91	1462	247	9.10	438	2.19	2.35	0.23	69.8	0	0	21.9	
11/1 19:00	2.3	1.53	84.9	289	145	161	0.54	4.18	878	6.32	1399	430	7.46	669	1.81	2.35	0.24	52.1	11	0	20.0	
11/1 22:50	1.3	1.53	45.2	265	145	147	1.04	3.46	897	6.72	1474	286	7.89	459	2.17	2.29	0.23	68.8	0	0	(23.4)	
11/2 00:50	2.5	1.61	44.4	260	142	144	1.07	3.46	925	6.65	1511	285	7.80	455	2.05	2.25	0.23	70.0	0	0	22.0	
11/2 08:30	1.8	1.61	94.3	293	146	186	0.57	4.15	936	5.98	1454	582	7.35	897	1.51	2.27	0.23	38.2	27	0	20.0	
11/2 14:49	1.2	1.61	64.7	275	146	175	0.67	3.71	885	6.94	1480	308	8.24	508	1.98	2.25	0.23	65.7	0	0	24.7	
11/12 06:00	3.1	2.22	42.8	264	148	146	0.83	3.56	1485	7.91	2693	566	9.08	1000	2.21	1.85	0.23	62.9	0	0	(22.9)	
11/12 11:08	1.8	2.22	44.1	267	148	146	0.78	3.84	1413	7.93	2566	428	8.99	752	2.61	2.40	0.23	70.7	0	0	21.7	
11/12 13:53	2.6	2.22	44.1	269	148	146	0.80	3.98	1285	7.84	2317	370	8.92	646	2.32	2.36	0.23	72.2	0	0	20.8	
11/12 20:29	1.8	2.22	44.2	272	148	146	0.80	3.98	1191	8.02	2181	328	9.12	583	2.44	2.61	0.23	73.3	0	0	(23.0)	
11/13 00:00	5.3	1.68	44.9	269	147	146	0.81	3.77	1120	8.44	2128	437	9.27	786	2.60	2.22	0.23	63.3	0	0	23.6	
11/13 10:14	3.3	1.68	93.8	299	147	145	0.52	4.62	1437	6.05	2244	908	7.33	1399	1.95	2.01	0.23	37.8	25	0	19.8	
11/14 16:30	2.8	2.57	93.0	308	149	146	0.47	4.95	1520	7.85	2741	1025	8.75	1764	1.41	1.98	0.23	35.5	35	0	(26.9)	
11/28 04:21	1.8	1.78	43.6	264	151	149	1.11	3.42	977	7.33	1686	560	8.47	942	1.60	1.64	0.05	44.2	0	0	(24.3)	
11/30 13:46	2.4	2.79	52.4	269	156	155	0.90	3.75	1240	6.47	1999	631	7.57	989	1.80	1.61	0.19	50.6	0	0	24.1	
11/30 18:35	4.0	2.79	90.9	291	151	149	0.64	3.23	1614	7.99	2946	1058	9.21	1892	0.98	0.96	0.20	36.0	0	0	31.5	
12/01 02:05	5.0	2.88	52.6	268	149	147	0.86	4.01	1798	6.79	2972	1377	7.75	2188	0.69	0.92	0.16	26.5	25	0	26.8	
12/01 11:16	1.7	2.88	92.4	288	157	196	0.71	3.62	1731	7.68	3078	1019	8.87	1770	1.05	0.98	0.17	42.6	0	0	66.8	
12/01 19:06	1.2	2.88	92.8	286	158	157	0.74	3.56	1822	6.63	2975	1464	8.04	2380	0.70	0.97	0.00	19.8	30	0	49.05	
12/04 15:27	4.0	2.36	93.0	293	150	149	0.66	3.92	1829	6.44	2943	1357	7.91	2184	0.71	0.98	0.00	26.5	28	0	52.71	
12/04 21:37	1.7	2.36	87.9	283	152	150	0.75	3.60	1830	6.25	2900	1568	7.42	2456	0.89	1.02	0.00	16.3	29	0	47.32	
12/05 01:07	0.8	2.77	43.3	256	147	146	1.12	3.05	1635	8.20	3040	1220	9.05	2152	1.11	1.03	0.00	15.4	26	0	41.51	
																			29.2	0	0	(27.5)

**PROCESS RUN DATA - HYDRATED LIME A**

Test Date	Test Duration (hrs)	Coal S (wt%)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	Water/Fl Gas (gal/ton)	Gas Analyzer Data				Ca/S, mol	Nu/Ca (mol)	% SO <sub>2</sub> Form-oval	By-pass %	Recycle Feed, Mtpd/hr	Aprx to Adiab Satur °F				
				Humidifier		ESP			Humidifier Inlet		ESP Outlet								Sys Hum			
				Inlet	Out-let	Inlet	TV	SO <sub>2</sub> ppm	O <sub>2</sub> %	SO <sub>2</sub> (corr)	SO <sub>2</sub> ppm	O <sub>2</sub> %	SO <sub>2</sub> (corr)									
12/05 02:37	1.2	2.77	80.0	279	151	149	177	3.59	1597	7.62	2825	1368	7.78	2175	103	1.24	0.00	30.9	22	4100	(28.1)	
12/05 04:27	2.5	2.77	43.6	253	147	147	145	2.97	1632	8.17	3026	1177	9.15	2093	1.09	1.03	0.60	30.7	0	6490	21.8	
12/05 11:58	2.0	2.77	92.2	290	150	149	190	3.93	1807	6.44	2905	1358	7.52	2122	0.97	1.28	0.20	27.2	28	5745	(26.0)	
12/05 16:38	4.5	2.77	90.7	288	151	149	190	3.91	1827	6.36	2919	1329	7.62	2092	1.12	1.46	0.20	28.4	28	0	24.5	
12/06 01:58	4.0	2.72	44.0	261	149	146	145	3.33	1611	7.99	2942	847	9.21	1513	1.62	1.49	0.20	48.6	0	0	23.4	
12/06 08:48	3.3	2.72	91.2	295	151	149	191	4.09	1735	6.54	2811	1308	7.35	2018	1.10	1.50	0.20	28.2	28	0	25.0	
12/11 20:11	2.0	2.57	81.5	283	149	148	165	0.65	3.90	2055	6.11	3222	1451	7.38	2242	1.41	1.34	0.09	30.4	12	3451	23.0
12/12 02:11	3.3	3.18	47.4	262	145	144	142	0.94	3.46	1877	7.89	3396	1066	8.70	1827	1.36	1.31	0.18	46.1	0	2481	21.1
12/14 04:15	1.7	2.86	44.1	260	146	145	144	0.98	3.32	1552	8.68	3016	1095	9.64	2032	1.62	1.60	0.00	32.8	0	0	23.3
12/14 10:56	1.8	2.86	93.6	285	150	150	184	0.69	3.78	1738	6.44	2793	1437	7.39	2222	1.50	2.17	0.00	20.3	25	0	24.0
12/14 18:46	1.2	2.86	92.9	283	151	150	183	0.71	3.71	1899	6.50	3069	1610	7.35	2484	1.49	1.96	0.00	19.5	25	0	(26.4)
01/05 22:47	1.3	1.61	41.7	266	160	159	157	2.55	1106	8.11	2042	947	9.45	1730	1.51	1.03	0.00	14.8	0	0	(35.8)	
01/06 23:53	1.1	1.60	41.2	262	159	159	156	1.18	2.45	1139	8.05	2091	981	9.42	1785	1.60	1.00	0.00	14.6	0	0	(34.9)
01/08 21:45	1.1	1.60	88.8	296	152	152	192	0.64	3.11	1180	6.00	1836	1039	7.80	1657	0.74	0.69	0.00	9.7	28	0	(27.1)
01/09 01:25	1.5	1.39	45.9	276	170	170	171	1.98	968	7.71	1725	785	9.04	1384	1.63	0.90	0.00	19.8	0	0	(48.3)	
01/09 03:26	1.3	1.39	45.9	273	170	170	168	1.35	1.84	1024	6.67	1704	814	8.46	1368	1.62	0.82	0.00	19.7	0	0	(44.2)
01/09 05:27	1.5	1.39	60.5	282	171	170	182	1.10	2.08	1051	6.22	1663	907	7.91	1459	1.18	0.73	0.00	11.7	10	0	(46.2)
01/09 22:33	1.5	1.39	44.2	275	148	148	145	1.00	3.49	928	7.77	1662	667	9.22	1194	1.44	1.13	0.00	30.2	0	0	22.9

**PROCESS RUN DATA - HYDRATED LIME G**

Test Date	Test Duration (hrs)	Coal S (%)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	Fl Gas (ppm)	Gas Analyzer Data				Na/Ca (mol)	% SO2 Rem-oval	By-pass %	Recycle Feed, kbb/hr	Appr. to Adiab. Satur. °F						
				Inlet	Out-	Humidifier			Humidifier Inlet	ESP Inlet	ESP Outlet	SO2 ppm						O2 %	SO2 (corr)	SO2 (corr)			
01/16	16 09	1.35	105	312	154	151	199	4.62	924	6.24	1463	874	7.33	1347	0.82	1.10	0.00	8.2	29	0	31.8		
01/16	19 29	1.35	107	300	151	148	192	4.40	1055	5.81	1617	991	7.02	1492	0.81	1.04	0.00	7.7	28	0	25.6		
01/16	22 29	1.4	43	279	151	149	149	0.93	885	8.35	1668	661	9.78	1242	1.37	1.14	0.00	24.0	0	0	26.5		
01/17	00 00	5.5	44	274	146	145	144	0.93	832	8.34	1568	632	9.71	1181	1.45	1.20	0.00	24.8	0	0	(22.6)		
01/17	09 00	1.2	91	297	158	156	173	0.66	407	9.79	5.15	1433	819	6.48	1187	1.97	2.00	0.00	17.2	11	0	32.2	
01/17	17 51	1.0	103	309	160	157	209	0.64	4.38	9.71	5.65	1472	916	7.04	1381	1.39	1.82	0.00	6.2	33	0	33.0	
01/17	20 21	1.3	103	301	159	156	207	0.69	4.19	10.08	5.67	1531	937	7.25	1435	1.42	1.82	0.00	6.0	34	0	(33.5)	
01/17	22 41	1.0	43	279	157	155	154	0.94	3.54	8.11	8.85	1601	632	10.2	1230	2.35	1.90	0.00	24.5	0	0	32.0	
01/18	00 41	5.2	43	272	147	145	143	0.93	3.78	8.34	8.72	1626	590	10.0	1131	2.16	1.82	0.00	30.2	0	0	21.3	
01/18	11 41	1.2	86	292	161	159	180	0.72	3.82	9.83	6.50	1589	915	7.32	1408	0.81	0.82	0.00	11.4	15	0	36.1	
01/18	15 31	1.0	67	281	155	153	156	0.75	3.69	9.30	7.25	1594	811	8.22	1336	0.95	0.83	0.00	16.2	1	0	(31.5)	
01/18	18 02	3.8	97	294	157	156	202	0.73	3.44	10.00	6.52	1617	959	7.51	1497	0.78	0.98	0.00	8.0	33	0	(32.2)	
01/19	01 32	1.3	43	261	146	145	143	0.93	2.94	7.62	9.77	1647	593	10.7	1214	1.37	1.08	0.00	26.3	0	0	(25.2)	
01/19	04 32	1.3	66	272	147	146	165	0.86	3.21	8.90	7.60	1555	555	10.3	1095	1.40	1.19	0.00	28.2	0	0	(23.7)	
01/19	13 26	1.0	66	272	147	146	165	0.86	3.21	8.90	7.60	1555	555	10.3	1095	1.40	1.19	0.00	28.2	0	0	(23.7)	
01/19	19 27	2.5	102	291	150	148	198	0.67	3.64	10.49	6.29	1667	916	7.25	1402	1.76	2.08	0.00	16.0	34	0	(24.5)	
01/20	00 17	1.7	44	266	144	143	144	0.88	3.15	7.63	9.62	1624	541	10.6	1094	1.43	1.15	0.18	32.5	0	0	(25.0)	
01/20	03 17	1.0	43	266	144	143	142	0.88	3.16	7.62	9.59	1618	497	10.7	1016	1.42	1.15	0.22	36.9	0	0	(23.7)	
01/20	05 27	4.5	43	268	144	144	143	0.89	3.17	7.09	9.36	1471	436	10.5	875	1.44	1.28	0.24	40.2	0	0	23.1	
01/20	17 57	1.0	108	288	151	150	202	0.66	3.54	9.98	5.19	1464	897	6.68	1319	0.86	1.23	0.23	9.9	37	0	28.0	
01/20	19 17	1.0	109	292	155	155	207	0.69	3.48	10.13	5.32	1499	930	6.77	1375	0.86	1.21	0.24	8.6	38	0	(28.8)	
01/20	21 27	0.8	64	280	155	155	165	0.85	3.16	9.34	6.44	1503	723	7.96	1167	1.27	1.20	0.24	22.3	8	0	33.2	
01/20	22 47	1.0	43	268	155	155	153	1.03	2.89	7.93	8.76	1552	560	10.1	1078	1.53	1.28	0.23	30.2	0	0	(32.8)	
01/21	03 27	2.3	43	261	146	146	144	1.08	2.98	8.28	8.22	1543	549	9.59	1014	1.36	1.11	0.23	34.2	0	0	(22.5)	
01/21	12 28	1.0	42	262	146	146	143	0.96	2.97	7.39	9.04	1485	508	10.2	991	1.32	1.15	0.23	33.4	0	0	(23.4)	
01/22	03 28	2.3	43	265	144	144	142	0.97	3.16	6.54	8.43	1241	403	9.84	762	1.26	1.43	0.21	38.8	0	0	24.2	
01/22	08 38	3.0	107	295	150	149	203	0.66	3.69	9.87	5.38	1468	882	6.98	1325	0.77	1.17	0.25	9.3	36	0	(24.4)	
01/22	15 58	3.0	105	292	149	148	201	0.64	3.71	10.36	5.32	1534	793	6.92	1186	1.54	2.22	0.26	22.6	36	0	26.4	
01/22	20 28	1.0	105	292	149	148	199	0.65	3.73	10.81	5.49	1619	792	7.72	1257	1.52	2.02	0.26	22.4	35	0	25.5	
01/23	00 28	2.3	45	265	149	148	146	0.97	3.07	10.17	6.89	1695	443	8.83	768	2.44	1.97	0.25	54.7	0	0	(22.3)	
01/23	04 08	6.4	106	295	154	153	204	1.06	2.96	9.79	7.28	1683	597	8.86	1036	2.49	2.02	0.00	38.1	0	0	23.9	
01/23	11 29	1.0	106	295	154	153	204	0.69	3.49	10.62	5.16	1554	940	6.71	1384	1.56	2.14	0.00	11.0	36	0	(27.6)	
01/23	18 49	1.7	108	294	150	149	201	0.63	4.12	10.34	4.94	1491	841	6.45	1216	1.41	1.90	0.26	18.3	35	0	25.5	
01/24	00 00	6.0	162	44	267	146	145	143	0.95	3.51	8.23	8.20	1531	380	9.31	686	1.86	2.00	0.24	55.0	0	0	22.8
01/24	10 29	1.0	106	294	150	150	201	0.66	3.92	9.27	5.69	1408	822	6.88	1225	1.12	2.03	0.00	13.0	36	0	(24.4)	
01/24	18 24	2.5	103	298	153	154	205	0.63	4.07	16.71	5.99	2598	1521	7.18	2316	1.00	0.96	0.76	10.8	36	0	26.2	

**PROCESS RUN DATA - HYDRATED LIME G**

Test Date	Test Duration (hr)	Coal S (wt %)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	Water/Fl Gas (gal/klb)	Gas Analyzer Data			Ca/S, mol	Na/Ca (mol)	% SO2 Rem-oval	By-pass %	Recycle Feed, klb/hr	Appr in Ashab Satur = F						
				Humidifier Inlet	Humidifier Out-let	ESP Inlet			Humidifier SO2 ppm	O2 %	SO2 %							Sys Hum	ESP Outlet SO2 ppm	O2 %	SO2 %		
01/25	00:54	5.0	2.58	43	270	150	149	147	0.88	3.45	1670	7.67	2965	1231	8.75	2119	1.02	0.89	0.25	28.6	0	0	26.2
01/25	05:34	1.6	2.58	106	300	161	160	213	0.62	3.79	1720	5.81	2639	1685	6.99	2532	0.62	0.93	0.00	3.7	37	0	34.6
01/25	16:44	1.8	2.58	104	303	160	160	213	0.58	4.02	1856	6.13	2914	1559	7.38	2410	0.99	1.38	0.25	17.3	37	0	34.3
01/25	19:54	2.7	2.58	105	293	160	160	208	0.62	3.73	1874	6.12	2941	1586	7.25	2429	0.99	1.37	0.20	17.4	36	0	33.7
01/26	01:54	1.0	2.71	43	267	147	145	144	0.77	3.50	1588	8.47	3026	884	9.82	1667	1.57	1.37	0.20	45.0	0	0	23.7
01/26	10:04	3.7	2.71	103	293	160	160	209	0.61	3.76	1890	6.07	2956	1541	7.37	2379	0.93	1.36	0.20	18.9	37	0	36.2
01/26	20:35	1.2	2.71	101	292	163	165	213	0.69	3.53	2007	6.23	3177	1795	7.56	2811	0.68	0.92	0.20	11.5	38	0	(38.4)
01/27	02:15	6.0	2.71	44	258	148	148	146	0.82	3.27	1750	8.36	3302	864	9.87	1636	1.62	1.32	0.11	50.1	0	0	(24.8)
01/27	10:45	4.2	2.71	44	263	149	149	147	0.83	3.18	1645	8.56	3160	1152	10.0	2210	1.62	1.39	0.00	30.1	0	0	24.7
01/27	17:26	3.2	2.71	44	270	167	170	167	1.04	2.52	1570	8.73	3064	1336	10.2	2603	0.60	0.52	0.00	15.0	0	0	(46.2)
01/29	12:24	3.3	2.71	98	285	155	154	203	0.69	3.60	1937	6.57	3263	1893	7.82	3025	0.64	0.86	0.00	7.5	37	0	29.6
02/01	22:52	1.3	2.90	41	260	152	150	149	0.93	3.11	1646	8.68	3199	1236	9.68	2304	1.45	1.33	0.00	28.0	0	0	(31.3)
02/02	00:00	3.7	2.41	42	257	152	149	149	0.95	3.06	1612	8.63	3118	1220	9.66	2269	1.69	1.36	0.00	27.5	0	0	27.6
02/02	19:57	1.8	2.41	102	286	157	155	205	0.68	3.70	1372	6.14	2156	887	7.36	1369	0.78	1.38	0.18	36.8	38	0	(33.5)
02/03	03:28	4.8	1.54	42	255	151	150	149	0.92	3.05	875	8.82	1722	433	10.0	831	2.00	1.81	0.18	51.0	0	0	(30.9)
02/03	18:19	1.1	1.54	106	284	160	160	201	0.73	3.44	1025	5.43	1530	471	9.89	895	1.98	1.78	0.18	48.8	0	0	(30.7)
02/05	18:21	1.6	1.31	100	288	153	153	197	0.62	3.74	1057	5.63	1600	899	7.06	1357	0.65	0.78	0.18	15.2	32	35.78	29.3
02/06	22:58	1.1	1.49	65	266	151	150	149	0.72	3.33	982	6.81	1626	726	7.86	1165	0.77	0.75	0.18	29.3	0	9108	26.1
02/06	00:00	4.6	1.46	43	260	147	146	144	0.93	3.29	913	8.26	1706	638	9.55	1176	0.84	0.74	0.20	30.9	0	10346	24.9
02/07	08:52	1.7	1.46	65	263	151	151	149	0.76	3.19	1041	6.73	1712	638	9.61	1180	0.89	0.76	0.19	31.1	0	9611	25.0
02/07	15:52	4.2	1.46	65	268	151	149	148	0.71	3.36	976	6.75	1607	711	8.06	1157	0.75	0.71	0.17	27.7	0	8442	26.1
02/08	00:03	5.8	1.39	45	260	146	145	144	0.96	3.31	936	7.38	1622	643	8.71	1103	0.84	0.72	0.18	32.0	0	9159	(22.8)
02/08	08:23	1.0	1.39	65	265	151	150	149	0.76	3.22	980	6.11	1536	753	7.52	1176	0.79	0.72	0.18	23.4	0	7607	27.8
02/08	10:23	8.5	1.39	66	272	149	147	147	0.67	3.49	946	6.91	1578	707	8.12	1157	0.79	0.72	0.17	26.5	0	9119	(25.3)
02/08	22:33	1.4	1.39	45	265	148	147	147	0.90	3.31	903	8.38	1706	710	9.57	1310	0.83	0.69	0.18	23.6	0	7222	27.9
02/09	00:24	4.3	1.41	46	262	146	145	144	0.90	3.35	931	8.20	1733	697	9.43	1271	0.82	0.68	0.17	26.8	0	7672	25.2
02/09	10:04	2.7	1.41	64	271	149	146	148	0.65	3.53	920	7.48	1608	710	8.81	1227	0.76	0.70	0.18	24.0	0	8638	26.7
02/09	13:44	1.2	1.41	64	272	153	151	152	0.68	3.42	884	7.71	1575	707	8.86	1281	0.77	0.75	0.17	22.1	0	9347	30.6
02/09	20:25	2.5	1.41	64	267	151	150	150	0.72	3.33	1007	7.14	1712	820	8.48	1381	0.75	0.64	0.17	19.4	0	9663	28.8
02/10	02:05	8.3	1.44	45	253	146	145	143	1.02	3.13	999	8.04	1832	803	9.10	1422	0.76	0.62	0.18	22.9	0	7609	24.4
02/11	05:36	7.2	1.26	43	259	144	144	142	0.94	3.16	823	8.77	1612	645	9.93	1228	0.93	0.75	0.00	23.6	0	9774	25.1
02/12	00:56	1.8	1.24	45	259	142	143	142	0.89	3.17	729	8.84	1437	600	9.86	1135	1.17	1.07	0.00	22.1	0	8011	25.1
02/12	03:46	1.0	1.24	45	259	143	143	141	0.88	3.20	710	8.95	1416	525	9.96	1002	0.94	0.88	0.00	29.2	0	7080	24.5
02/12	15:47	1.2	1.24	43	260	153	152	151	0.96	2.90	858	9.29	1765	754	9.97	1441	0.94	0.71	0.00	18.7	0	7931	32.1

**PROCESS RUN DATA - HYDRATED LIME G**

Test Date	Test Duration (hrs)	Coal S (wt %)	Gross MW	Temperatures, °F			Air/Water (lb/lb)	Water/FI Gas (gal/lb)	Gas Analyzer Data						Ca/S, mol	Na/Ca (mol)	% SO2 Rem-oval	By-pass %	Recycle Feed, lb/hr	Appr 10 Adiab Satur °F
				Humidifier Inlet	Humidifier Out-let	ESP Inlet			SO2 ppm	O2 %	SO2 ppm	O2 %	SO2 ppm	SO2 %						
02/12 21:37	1.0	1.24	100	287	155	196	0.59	3.70	1119	5.87	1724	1050	7.26	1610	0.63	0.65	6.9	31	6893	31.3
02/13 00:47	4.2	1.36	45	259	147	145	0.86	3.19	879	8.57	1690	701	9.69	1307	0.88	0.74	22.7	0	8581	26.3
02/13 11:37	2.2	1.36	102	292	171	208	0.64	3.38	1103	5.42	1646	1046	7.25	1601	0.61	0.71	2.7	31	4991	(46.1)
02/13 15:07	2.2	1.36	102	296	166	207	0.60	3.62	1139	5.35	1689	1097	7.12	1664	0.61	0.70	1.3	31	4814	42.8
02/13 18:58	1.8	1.36	102	289	166	165	0.71	3.40	1137	5.72	1733	1101	7.27	1689	0.59	0.70	2.7	37	4490	41.2
02/14 01:28	2.3	1.27	47	250	147	146	0.92	2.94	793	9.69	1701	692	10.2	1354	0.89	0.77	20.3	0	7697	27.6
02/14 04:18	1.7	1.27	45	249	146	144	0.91	2.97	739	10.1	1654	616	10.6	1244	0.88	0.77	24.7	0	8730	28.5
02/14 09:38	4.5	1.27	79	270	151	150	0.79	3.36	1048	6.23	1658	1002	7.33	1544	0.66	0.70	20.0	24.7	3853	30.3
02/15 00:00	1.2	1.25	46	252	146	145	0.90	3.06	846	9.05	1702	700	9.79	1317	0.94	0.75	22.7	0	7756	25.9
02/15 03:42	4.2	1.25	47	256	144	143	0.86	3.18	826	9.04	1661	652	9.74	1221	0.91	0.74	26.3	0	8280	24.1
02/15 12:51	1.8	1.25	75	279	149	148	0.71	3.60	929	7.50	1627	906	8.36	1509	0.69	0.74	7.3	27	4199	(28.4)
02/15 18:32	1.3	1.25	96	282	147	145	0.86	3.74	1025	6.21	1619	985	7.14	1495	0.63	0.72	7.7	33	4997	(23.9)
02/16 00:00	2.0	1.24	46	255	143	142	0.86	3.13	825	8.64	1597	658	9.67	1225	0.97	0.78	23.3	0	7165	24.0
02/16 03:31	1.5	1.24	46	257	135	133	0.77	3.49	815	8.72	1580	516	9.64	958	0.95	0.78	39.7	0	7361	16.6
02/16 07:22	2.2	1.24	46	258	136	133	0.79	3.38	831	8.74	1623	683	9.61	1264	0.00	0.00	21.7	0	6827	19.2

DEPOSIT SAMPLES

ANALY. NUMBER	SAMPLE DATE	SAMPLE NUMBER	SAMPLE DESCRIPTION	MOIS-TURE	ASH	C	H	N	S	ASHELEMENTALS (DRY, AS RECEIVED)											SULFUR FORM		CO3 wt %
										SiO2	Al2O3	TiO2	Fe2O3	CaO	MgO	Na2O	K2O	P2O5	SO3	SULFITE	SULFATE		
895022	10/09/89	100989	FLOOR DEPOSIT	0.3	87.88	1.86	0.37		13.87	8.31	0.31	7.44	2.28	0.32	29.56	0.64					2.05	11.63	0.07
895214	10/09/89	100989-1	FLOOR DEPOSIT	0.4	87.82	5.63	0.50		31.48	14.11	0.67	22.13	11.73	0.61	3.80	1.33							0.47
895215	10/09/89	100989-2	TV DEPOSIT	1.2	77.82	4.75	0.80		5.38	2.49	0.13	2.57	41.61	0.50	3.30	0.24							29.76
895216	10/09/89	100989-3	DEPOSIT AFTER TV	11.7	83.87	1.39	0.08		20.87	9.48	0.42	10.00	30.85	0.64	2.80	0.91							15.94
895774	11/17/89	EDGEWATER 1	NOZZLE TULIP	0.1	84.45	3.86	0.35	0.07	14.31	7.28	0.35	6.56	1.00	0.29	27.09	0.69							
895775	11/17/89	EDGEWATER 2	NOZZLE SHIELD DEP	0.5	81.37	2.02	0.35	0.08	12.58	6.60	0.32	7.49	6.14	0.39	27.61	0.49							
895778	11/17/89	EDGEWATER 3	AIRFOIL DEPOSIT	0.7	88.24	1.51	1.39	0.16	35.44	17.10	0.78	18.70	11.76	0.75	1.28	1.53							
898165	12/08/89	120889	TV DEPOSIT	0.4	86.86	1.08	0.22	0.19	9.29	4.49	0.19	6.47	46.05	0.63	1.36	0.33					4.65	1.23	3.73
898166	12/15/89	121589-1A	ESP 1A WIRE DEPOS	0.3	87.12	0.47	0.08	0.20	8.86	4.72	0.23	4.27	28.82	0.50	4.14	0.40					0.03	17.13	0.17
898167	12/15/89	121589-2A	ESP 2A WIRE DEPOS	0.3	85.51	1.08	0.22	0.19	15.15	7.43	0.35	6.78	26.75	0.56	3.53	0.69					2.94	10.64	0.26
901263	02/16/90	0216901800	TV DEPOSIT	0.4	86.53	1.44	0.88	0.08	10.35	5.22	0.26	3.42	28.21	0.99	6.12	0.64					10.63	3.88	1.81

COAL SAMPLES

ANALY. NUMBER	SAMPLE DATE	SAMPLE NUMBER	MOIS-TURE	VM	ASH	C	H	N	SULFUR FORMS TOTAL	BTU	ASH ELEMENTALS (DRY BASIS)										
											VM	ASH	C	H	N	Pyritic	Sulfate	ASHING TEMP	SiO2	Al2O3	TiO2
894920	10/02/99	100209-2440	3.7		11.84	71.81	4.82	1.38	2.16		750	48.95	22.84	1.06	19.01	1.61	0.73	0.16	2.20	0.35	1.86
900908	01/26/90	0126901735	4.3		12.67	70.97	4.78	1.21	2.77	12748	750	46.00	23.06	1.01	23.24	1.31	0.86	0.49	2.24	0.38	1.41
900909	01/26/90	0126901730	4.0		12.84	71.12	4.76	1.24	3.00	12735	750	44.72	22.74	0.98	24.07	1.42	0.87	0.55	2.23	0.38	1.64
900950	02/08/90	020890-0100C	4.4	34.84	10.38	74.55	4.95	1.39	1.46	0.93	750	50.04	23.35	1.13	17.59	1.27	0.83	0.64	2.33	0.34	1.16
900951	02/10/90	021090-0540C	4.1	34.13	11.17	73.87	4.93	1.37	1.66	0.93	750	49.01	22.59	1.08	19.36	1.42	0.90	0.50	2.36	0.34	1.29
900952	02/10/90	021090-1920C	3.9	34.38	10.33	74.78	4.89	1.35	1.44	0.87	750	50.73	23.70	1.12	16.53	1.45	0.90	0.38	2.45	0.38	1.28
901162	02/12/90	021290-2160	4.3	35.00	11.00	74.09	4.93	1.40	1.34	0.66	750	53.34	23.91	1.05	14.46	1.59	0.95	0.38	2.48	0.32	1.48
901183	02/10/90	021090-1920	4.1	34.98	10.31	74.48	4.92	1.43	1.43	0.83	750	51.64	24.04	1.06	14.62	1.37	0.89	0.38	2.44	0.36	1.16
901420	01/26/90	0126902250	4.1		13.62	70.07	5.09	1.31	3.28	1.93	750	44.44	21.70	0.92	24.50	1.20	0.78	0.35	2.11	0.35	1.42

ESP HOPPER SAMPLES

ANALY. NUMBER	SAMPLE DATE	SAMPLE NUMBER	MOIS-TURE	ASH	C	H	N	S	Ashing Temp. °C	ASH ELEMENTALS (DRY BASIS)										SO <sub>3</sub>	SULFUR FORM SULFATE	CO <sub>2</sub> wt %	wt % Ca(OH) <sub>2</sub> Ash Anal	ion		
										SiO <sub>2</sub>	Al <sub>2</sub> O <sub>3</sub>	TiO <sub>2</sub>	Fe <sub>2</sub> O <sub>3</sub>	CaO	MgO	Na <sub>2</sub> O	K <sub>2</sub> O	P <sub>2</sub> O <sub>5</sub>	SULFITE							
894643	09/12/89	091289-1315	0.3	90.83	3.91	0.33	0.08	4.79	925	25.84	13.01	0.60	13.87	31.55	0.87	0.15	1.10	0.22	11.92	2.45	1.50	6.32	20.1	18.6		
894694	09/22/89	912289	0.1	88.55	3.46	1.13	0.08	3.90	As Rec	43.38	20.02	0.92	12.05	8.12	0.80	0.16	1.96	0.43	2.61	0.61	0.46	0.48	7.3			
894915	09/20/89	092089-1215	0.2	90.99	3.28	0.29	0.15	1.04	As Rec	24.15	12.15	0.58	6.45	28.78	0.73	3.04	1.06	0.16	9.99	2.72	1.55	1.00	31.0			
894916	09/22/89	092289-1650	0.5	88.93	3.56	1.04	0.16	4.00	As Rec	29.90	14.52	0.68	7.67	20.77	0.71	1.76	1.31	0.18	10.52	3.00	1.25	1.04	18.4			
894917	09/26/89	092689-1515	0.7	88.91	5.54	0.68	0.19	4.21	As Rec	23.29	11.93	0.55	7.94	27.13	0.63	2.12	1.04	0.17	12.84	3.63	1.33	1.42	24.5			
894918	10/04/89	100489-2025	0.9	88.68	5.55	0.80	0.15	5.13	As Rec	20.37	9.78	0.46	8.94	31.73	0.59	2.48	0.90	0.15	13.34	4.75	1.31	1.23	28.9			
894919	10/05/89	100589-1445	1.0	91.23	2.96	1.29	0.11	6.14	As Rec	21.78	10.47	0.47	9.26	29.13	0.62	3.32	1.01	0.15	13.16				30.4			
894946	10/06/89	100689	0.8	90.47	4.43	1.00	0.09	6.06	As Rec	17.52	8.89	0.40	8.58	32.54	0.65	3.61	0.77	0.13	11.26	5.31	1.56	0.79	30.6			
895020	10/11/89	1011891320	1.2	90.05	2.58	1.31	0.08	6.46	As Rec	16.82	7.72	0.34	8.16	35.64	0.70	4.07	0.84	0.06	19.90	6.26	0.81	1.36	34.5			
895150	10/18/89	1018891705	0.8	89.93	4.75	1.05	0.08	3.01	As Rec	23.72	12.08	0.58	6.45	26.31	0.68	2.98	0.97	0.20	7.53				29.4			
895213	10/27/89	1027890821	1.4	85.77	4.57	1.39	0.20	7.96	As Rec	14.06	7.07	0.31	6.81	33.57	0.61	4.58	0.64	0.06	19.90	6.61	1.28	1.36	33.7			
895613	12/01/89	120189-2330	0.7	89.24	2.89	1.03	0.14	4.32	As Rec	20.59	10.49	0.49	10.13	33.13	0.70	4.49	0.90	0.17	9.92				33.5			
895625	12/02/89	120289-2350	1.2	86.81	3.96	0.97	0.18	3.96	As Rec	21.18	11.05	0.48	11.15	36.60	0.71	3.34	0.96	0.17	10.81				33.2			
895922	12/04/89	1204890315	0.9	89.38	3.69	0.92	0.11	4.11	As Rec	22.22	11.60	0.46	9.94	33.06	0.64	0.20	0.90	0.14	9.91	3.22	0.62	1.02	33.2			
895923	12/05/89	1205890845	0.8	89.15	3.90	0.89	0.16	4.47	As Rec	22.01	11.29	0.44	11.58	31.92	0.68	0.20	0.93	0.18	10.28	3.24	0.60	0.96	31.4			
895924	12/06/89	1206890815	0.6	89.97	3.52	0.94	0.19	5.44	As Rec	21.01	10.66	0.44	11.30	31.06	0.65	1.48	0.86	0.17	11.18	3.43	0.86	1.48	30.5			
896162	12/11/89	1211890845	0.2	89.93	4.37	0.58	0.11	3.41	As Rec	19.38	10.18	0.45	10.35	28.43	0.67	3.62	0.81	0.14	14.06	3.94	1.17	1.04	27.9			
896164	12/12/89	1212890800	0.2	88.84	4.27	1.05	0.08	7.83	As Rec	16.70	14.15	0.41	16.59	19.39	0.73	1.65	1.15	0.18	18.84	2.27	0.82	0.85	18.6			
900058	11/15/89	111589-200 #1	0.9	88.98	4.22	1.05	0.09	7.57	As Rec	14.02	7.02	0.31	7.05	34.31	0.61	4.61	0.67	0.08	19.08	6.19	1.27	2.96	29.3			
900057	11/15/89	111589-0200 #2	1.0	88.98	4.22	1.02	0.09	7.57	As Rec	14.10	7.02	0.30	7.07	34.61	0.62	4.58	0.66	0.09	18.92	6.42	1.40	2.92	29.8			
900058	12/05/89	120589-0110	0.7	89.15	3.78	0.99	0.06	3.77	As Rec	14.98	7.17	0.31	6.86	34.81	0.62	4.68	0.66	0.10	18.17	6.07	1.40	2.64	31.2			
900059	12/05/89	120589-0110	0.7	89.15	3.78	0.99	0.06	3.77	As Rec	22.52	10.91	0.48	11.76	31.54	0.71	0.20	0.98	0.10	9.43	3.88	0.57	1.05	31.7			
900459	12/14/89	1214891200-1300	0.9	87.14	4.22	1.26	0.10	2.18	As Rec	22.32	10.64	0.43	12.75	33.88	0.67	0.16	0.84	0.17	5.38	1.75	0.19	0.83	38.6			
900492	01/17/90	011790-0830	0.8	89.90	5.48	0.68	0.06	1.78	As Rec	34.20	16.89	0.73	10.60	20.39	1.15	0.24	1.59	0.25	4.47	1.30	0.31	0.52	22.3			
900493	01/19/90	011990-0815	1.2	90.44	1.94	1.02	0.06	2.39	As Rec	29.87	13.63	0.57	7.77	31.84	1.26	0.16	1.28	0.19	5.96	1.71	0.38	0.69	35.5			
900494	01/19/90	011990-0815	0.8	91.10	2.24	0.87	0.04	2.07	As Rec	29.87	13.63	0.57	7.77	31.84	1.26	0.16	1.28	0.19	5.96	1.71	0.38	0.69	35.5			
900495	01/20/90	012090-0815	0.6	92.09	2.01	0.78	0.05	2.54	As Rec	31.18	14.55	0.68	7.79	25.44	1.13	0.22	1.38	0.19	5.18	1.37	0.42	1.05	31.1			
900641	01/21/90	012190-0825	0.6	87.82	6.13	0.87	0.12	4.03	As Rec	32.73	14.84	0.64	5.73	30.85	1.27	2.17	1.50	0.19	6.73	1.96	0.68	0.67	28.7			
900643	01/21/90	012190-0500	1.3	90.66	2.06	1.04	0.02	3.48	As Rec	27.40	12.81	0.55	9.88	28.79	1.27	3.52	1.23	0.12	9.83	2.99	0.57	0.87	24.2			
900644	01/23/90	012390-0415	1.0	90.83	2.91	0.91	0.03	3.48	As Rec	21.18	9.89	0.42	8.00	34.39	1.36	3.52	1.06	0.20	8.56	2.82	0.57	0.51	29.9			
900645	01/23/90	012390-0505	1.3	90.75	1.29	1.21	0.04	5.20	As Rec	18.37	8.88	0.38	8.22	34.63	1.31	5.04	0.89	0.09	12.67	2.94	0.96	0.73	38.2			
900646	01/27/90	012790-0820	1.4	90.22	4.75	0.66	0.08	2.72	As Rec	34.61	16.37	0.74	9.20	18.76	1.08	1.99	1.69	0.15	7.46	3.43	1.31	0.86	38.4			
900694	02/08/90	02089000410	1.0	90.22	4.75	0.66	0.08	2.72	As Rec	33.20	15.90	0.73	8.74	19.20	1.06	2.22	1.64	0.15	7.90	2.26	0.52	0.64	19.9			
900905	02/10/90	02109000410	1.0	90.22	4.75	0.66	0.08	2.72	As Rec	20.20	9.98	0.41	8.71	38.78	1.46	0.31	0.99	0.15	7.90	2.48	0.46	0.83	42.5			
900906	02/10/90	02109000530	1.8	88.64	1.64	1.48	0.03	3.16	As Rec	24.81	11.91	0.53	6.34	31.95	1.32	2.83	1.24	0.13	8.44	2.58	0.68	0.85	36.3			
900907	02/10/90	02109000500	1.2	89.10	2.93	1.23	0.05	3.38	As Rec	33.38	15.86	0.75	8.50	18.56	1.06	2.78	1.63	0.10	8.10	2.28	0.61	0.44	19.7			
901001	02/09/90	020990-0610	0.6	89.85	6.57	0.52	0.08	2.44	As Rec	35.37	16.93	0.78	10.16	14.70	0.98	2.08	1.72	0.23	5.74	1.86	0.58	0.40	14.4			
901002	02/09/90	020990-0400	0.6	89.85	6.57	0.52	0.08	2.44	As Rec	33.62	16.64	0.79	9.18	14.70	0.98	2.08	1.72	0.23	5.74	1.86	0.58	0.40	14.4			
901003	02/08/90	020890-1030	0.6	90.03	6.08	0.56	0.05	2.53	As Rec	33.76	15.78	0.74	9.18	14.70	0.98	2.08	1.72	0.23	5.74	1.86	0.58	0.40	14.4			
901004	02/10/90	021090-0430	0.4	88.16	7.26	0.49	0.02	2.28	As Rec	39.05	18.27	0.80	10.73	12.90	1.02	0.86	1.89	0.25	5.61	1.78	0.61	0.58	12.0			
901188	02/19/90	021990-0320	0.5	91.32	4.92	0.44	0.02	2.84	As Rec	37.65	17.39	0.77	10.40	15.32	1.07	1.48	1.82	0.23	5.74	1.90	0.54	0.81	17.2			
901190	02/13/90	021390-0500	0.5	91.82	4.08	0.49	0.02	2.28	As Rec	36.08	17.57	0.77	9.26	13.51	1.07	1.74	1.76	0.20	5.72	1.50	0.61	0.58	12.0			
901193	02/14/90	021490-0610	0.5	90.56	7.88	0.56	0.04	2.28	As Rec	38.83	17.37	0.77	9.26	13.51	1.07	1.74	1.76	0.20	5.72	1.50	0.61	0.58	12.0			
901196	02/15/90	021590-0620	0.7	88.16	7.88	0.56	0.04	2.																		

**RECYCLE ASH SAMPLES**

ANALY NUMBER	SAMPLE DATE	SAMPLE NUMBER	MOIS-TURE	ASH	C	H	N	S	ASH ELEMENTALS (DRY, AS RECEIVED)										SULFUR FORM		CO3 wt %	wt % Ca(OH)2	
									SiO2	Al2O3	TiO2	Fe2O3	CaO	MgO	Na2O	K2O	P2O5	SO3	SULFITE	SULFATE		Ash	Therm
895811	12/01/89	120189-1530	0.6	91.10	3.29	0.71	0.14	3.86	25.52	12.83	0.58	12.76	24.77	0.69	1.70	1.14	0.20	9.67	1.18	24.2			
895812	12/02/89	120289-1620	0.7	88.75	4.69	0.60	0.12	3.19	23.24	14.80	0.54	13.18	27.23	0.69	0.79	0.97	0.16	7.99	1.11	28.0			
895828	12/03/89	120389-1440	0.9					3.52	23.95	12.47	0.52	13.14	32.61	0.70	0.34	1.04	0.16	8.80		33.8			
895925	12/04/89	120489-0930	0.7	90.16	3.39	0.78	0.10	3.19	24.60	12.61	0.51	14.61	27.34	0.62	0.19	0.99	0.15	7.99	1.79	26.6			
895925	12/04/89	120489-1430	0.8	99.03	3.41	0.98	0.10	3.84	20.88	10.80	0.44	11.45	33.07	0.82	0.25	0.88	0.16	9.67	1.33	33.2			
895927	12/05/89	120589-0515	0.7	89.18	4.38	0.80	0.10	4.67	25.29	12.81	0.50	13.28	27.64	0.64	0.20	1.02	0.20	7.56	1.22	28.1			
896183	12/11/89	121189-1710	1.3	89.48	2.75	1.00	0.11	4.67	20.25	10.54	0.45	10.67	30.85	0.72	2.61	0.86	0.13	12.14	1.46	28.9			
900900	02/06/90	020690-0400	0.8	88.77	5.84	0.70	0.10	2.21	34.58	16.17	0.73	9.45	19.35	1.07	1.78	1.68	0.21	5.53	0.67	21.6	18.6		
900901	02/06/90	020690-1630	0.9	88.68	5.96	0.65	0.02	2.55	34.00	16.32	0.74	9.10	18.37	1.04	1.92	1.67	0.20	6.39	0.55	19.8	17.2		
900902	02/07/90	020790-0040	0.8	90.05	4.79	0.68	0.05	2.80	34.00	16.21	0.74	8.98	19.01	1.06	2.13	1.67	0.19	7.02	0.57	20.3	18.8		
900903	02/07/90	020790-1120	0.8	89.83	5.03	0.65	0.05	3.20	33.34	15.84	0.78	8.98	19.29	1.08	1.83	1.70	0.23	6.77	0.57	20.3	18.8		
900953	02/08/90	020890-0400R	0.6	90.58	4.27	0.68	0.02	2.03	37.81	17.36	0.78	8.98	19.29	1.08	2.80	1.85	0.20	7.99	0.56	19.6	16.5		
900954	02/08/90	020890-1440R	0.3	90.73	6.07	0.42	0.07	2.18	36.04	16.89	0.77	12.11	14.12	1.00	1.80	1.67	0.21	5.46	0.48	12.8			
900955	02/09/90	020990-0600R	0.4	92.18	5.58	0.51	0.03	1.95	37.08	17.45	0.60	14.74	11.30	0.96	1.56	1.69	0.18	4.88	0.31	14.8			
900956	02/09/90	020990-1300R	0.4	92.18	5.43	0.37	0.03	1.95	37.08	17.45	0.60	14.74	11.30	0.96	1.56	1.69	0.18	4.88	0.31	14.8			
900957	02/10/90	021090-0415R	0.5	88.72	6.94	0.56	0.05	2.77	31.89	15.27	0.71	11.48	16.65	1.03	2.13	1.54	0.20	6.93	0.44	11.7			
900958	02/10/90	021090-1915R	0.6	88.43	6.04	0.77	0.11	3.10	28.67	13.63	0.64	12.60	20.05	1.02	1.94	1.39	0.19	7.74	0.68	20.7			
901184	02/11/90	021190-0200	0.3	87.41	6.84	0.60	0.12	2.41	34.22	15.76	0.72	12.01	15.13	1.00	1.28	1.62	0.22	6.14	0.57	15.2	11.8		
901185	02/11/90	021190-0500	0.3	90.09	7.21	0.42	0.10	1.96	37.10	16.64	0.76	16.16	12.09	0.98	1.06	1.71	0.23	5.09	0.50	12.1	8.4		
901186	02/11/90	021190-1610	0.3	90.79	6.73	0.38	0.08	1.89	36.39	16.25	0.69	17.78	11.74	0.97	0.79	1.69	0.23	4.73	0.65	11.3			
901187	02/12/90	021290-0300	0.3	90.73	5.30	0.50	0.03	2.20	36.74	16.94	0.74	12.59	14.48	1.08	1.20	1.79	0.23	5.76	0.57	13.6	10.4		
901189	02/13/90	021390-0430	0.4	91.13	5.03	0.48	0.03	2.20	36.74	16.94	0.74	12.59	14.48	1.08	1.20	1.79	0.23	5.49	0.67	14.7	11.2		
901192	02/13/90	021390-2005	0.3	89.36	7.10	0.47	0.09	2.05	36.06	17.10	0.75	12.14	13.59	1.01	1.25	1.75	0.21	5.14	0.52	14.4	10.9		
901192	02/14/90	021490-0600	0.3	89.18	7.26	0.48	0.09	2.20	35.43	16.89	0.74	11.95	14.16	1.00	1.35	1.73	0.22	5.49	0.52	14.4	10.9		
901194	02/14/90	021490-1815	0.3	86.18	11.09	0.36	0.08	1.57	36.72	17.37	0.77	11.81	10.63	0.94	1.35	1.75	0.22	3.93	1.01	11.5	8.7		
901195	02/15/90	021590-0610	0.3	86.19	11.06	0.36	0.10	1.53	37.12	17.46	0.78	12.21	10.33	0.93	1.30	1.77	0.21	3.82	0.40	11.1	8.5		
901197	02/16/90	021690-0530	0.4	87.80	9.26	0.39	0.11	2.03	36.30	16.96	0.74	13.79	10.96	0.93	1.35	1.71	0.23	5.07	0.39	10.6	7.9		
901199	02/16/90	021690-0930	0.5	89.49	7.73	0.39	0.09	2.52	36.51	17.32	0.75	11.81	11.71	0.94	1.53	1.75	0.24	6.31	0.60	10.7	7.4		



## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
07/28/89	18:10	127.5	8.00		
07/31/89	18:55	130	8.38		
08/03/89	10:50	135	6.58		
08/03/89	12:30	136	6.74		
08/07/89	13:40	127.5	8.40		
08/08/89	18:00	127.5			
08/14/89	13:20	122.5	8.20		
08/14/89	14:00	127.5	8.32		
08/14/89	15:05	125.5	8.24		
08/14/89	16:55	124.5	7.43		
08/15/89	15:30	130	7.36		
08/16/89	10:55	125.5	8.58		
08/16/89	12:45	128	8.04		
08/16/89	15:00	126	7.88		
08/16/89	20:50	127	8.04		
08/17/89	19:35	126	7.12		
08/29/89	10:35	126	8.00		
08/29/89	15:40	126	7.72		
08/30/89	05:00	126	9.27		
09/01/89	12:40	129			
09/01/89	14:00	129			
09/06/89	12:45	122			
09/06/89	13:10	123	9.80		
09/06/89	14:30	125.5	7.60		
09/06/89	16:25	130.5			
09/06/89	19:40	128	7.60		
09/06/89	22:20	131			
09/07/89	02:15	128	7.91		
09/07/89	05:05	128.5	8.05		
09/07/89	20:30	128.5			
09/08/89	03:15	128.5	9.03		
09/08/89	15:30	128	8.50		
09/11/89	11:25	122	8.17	7.40	280
09/11/89	18:15	124	8.51	7.50	292
09/12/89	04:00	126	8.82	7.58	277
09/12/89	08:40	123.5	10.14	8.99	271
09/12/89	13:30	123	8.74	7.56	275
09/12/89	20:00	126.5	7.98	7.03	290
09/13/89	04:00	126	9.09	7.80	279
09/13/89	14:45	131.5			
09/13/89	15:35	129.5	8.91	7.73	282

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
09/13/89	15:40	127.5	8.91	7.73	282
09/13/89	18:00	128	8.70	7.47	273
09/13/89	21:15	128	7.84	6.68	280
09/14/89	10:10	127.5	8.12	7.10	286
09/14/89	15:20	128.5	8.09	7.12	281
09/14/89	19:18	123			263
09/15/89	11:15	128.5	8.22	7.33	284
09/15/89	12:35	129.5	8.43	7.60	286
09/15/89	15:00	124.5			
09/15/89	17:23	124	8.55	7.59	287
09/18/89	17:00	125.5	8.71	7.83	287
09/18/89	19:00	125	8.67	7.65	289
09/19/89	05:05	125.5	9.44	8.47	282
09/19/89	09:45	126	8.70	7.55	287
09/19/89	20:30	125.5	9.04	7.71	291
09/22/89	15:00	126	9.04	7.59	298
09/25/89	12:40	121.5	9.21	7.53	264
09/25/89	23:30	123	10.62	9.28	268
09/26/89	04:40	122	10.42	8.98	269
09/26/89	16:00	122.5	8.17	6.56	287
09/26/89	22:00	121	8.32	6.73	275
09/27/89	06:10	121.5	9.22	7.73	269
09/27/89	18:00	123	8.22	7.12	287
09/27/89	23:00	121.5	9.33	7.76	277
09/28/89	01:00	120	9.24	7.82	272
09/28/89	03:00	120.5	9.24	7.82	272
09/28/89	10:30	123.5			
09/28/89	15:35	125	9.05	8.00	290
10/02/89	21:30	124.5			
10/03/89	05:10	126	9.29	7.99	273
10/03/89	10:00	124	7.68	6.71	283
10/03/89	14:30	125	7.55	6.38	278
10/03/89	21:35	124.5	7.58		
10/04/89	05:00	122			
10/04/89	07:30	121.5			
10/04/89	10:00	123.5			
10/04/89	15:00	122	8.99	7.46	275
10/04/89	18:00	122			
10/04/89	22:00	120			
10/05/89	05:00	122	9.34	8.05	271
10/05/89	10:00	125.5	8.52	6.84	286

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
10/05/89	14:15	119	10.13	8.89	289
10/05/89	21:00	120.5	8.97	7.72	290
10/06/89	04:45	123	8.69	7.38	275
10/06/89	10:00	125.5			
10/06/89	14:00	121.5	8.60	7.54	292
10/06/89	19:43	121.5	9.17	7.64	290
10/07/89	10:00	119	9.62	8.44	265
10/10/89	13:15	121	9.51	7.79	282
10/10/89	16:00	123.5	9.14	7.57	280
10/10/89	20:30	124	8.51	7.07	277
10/11/89	03:00	120	9.80	8.40	272
10/11/89	05:00	121	9.60	8.40	265
10/11/89	09:00	118	11.37	10.40	262
10/11/89	21:00	122	9.18	7.67	274
10/12/89	05:00	121	9.78	8.16	271
10/12/89	11:45	123	9.39	7.92	280
10/12/89	16:15	123	8.83	7.45	281
10/13/89	12:30	123	8.69	7.21	286
10/13/89	15:00	123.5	8.46	6.99	289
10/16/89	18:15	123	9.22	7.65	284
10/16/89	20:15	123.5	9.19	7.60	271
10/16/89	22:25	123.5	9.20	7.69	274
10/17/89	05:00	121	9.10	7.53	269
10/17/89	10:00	121	8.25	6.73	271
10/17/89	15:00	123	8.97	7.26	265
10/17/89	19:00	123			
10/18/89	05:00	124	9.66	8.13	264
10/18/89	09:00	123		6.11	286
10/18/89	20:20	123	7.91	6.52	282
10/18/89	22:30	124			
10/19/89	05:00	125	8.13	6.73	251
10/19/89	09:15	125	7.60	6.26	273
10/19/89	14:00	125.5	7.90	6.18	295
10/25/89	18:20	126		5.97	312
10/26/89	05:30	121		7.84	273
10/26/89	09:00	125		5.57	303
10/26/89	13:00	126.5		5.58	309
10/26/89	15:32	126.5		6.04	305
10/26/89	17:50	127		5.54	307
10/26/89	21:20	126		5.50	310
10/27/89	03:45	122		7.94	274

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
10/31/89	02:40	124		7.62	269
10/31/89	05:00	124		7.59	269
10/31/89	23:10	119.5		7.74	265
11/01/89	04:50	120		7.64	265
11/01/89	08:25	125		6.52	293
11/01/89	13:00	125		6.56	270
11/01/89	15:00	122		7.88	258
11/01/89	17:00	122		7.35	261
11/01/89	21:15	125		6.38	290
11/02/89	00:40	122		6.57	261
11/02/89	05:30	122		6.29	257
11/02/89	07:30	126		5.90	289
11/02/89	09:00	126		5.93	293
11/02/89	13:00	125		6.09	292
11/02/89	16:10	125		6.89	274
11/12/89	04:00	122		8.09	265
11/12/89	11:20	123		7.75	266
11/12/89	15:00	124		7.86	269
11/12/89	19:15	125		6.83	292
11/13/89	05:00	121		8.36	268
11/13/89	11:15	125.5		5.87	298
11/14/89	09:45	126		7.65	292
11/14/89	14:30	124			
11/14/89	23:00	125.5		7.45	270
11/15/89	05:20	125		8.05	269
11/28/89	02:20	123		7.31	265
11/28/89	14:15	124		6.57	278
11/29/89	09:07	120		6.86	292
11/29/89	20:00	122.5		7.65	298
11/30/89	04:50	123		8.30	283
11/30/89	12:35	121.5		7.33	301
11/30/89	15:30	121.5		8.13	266
11/30/89	19:30	122.5		7.15	292
12/01/89	07:00	124		7.51	268
12/01/89	09:15	129		5.67	291
12/01/89	20:00	125.5		6.47	287
12/02/89	03:00	124		8.12	263
12/02/89	16:00	126		6.60	298
12/02/89	20:20	125.5		6.28	295
12/03/89	01:00	124		7.98	260
12/03/89	05:00	125		7.98	260

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
12/03/89	09:00	120		8.10	258
12/03/89	15:00	120		7.80	259
12/03/89	17:15	125			276
12/03/89	18:45	125			
12/04/89	03:00	121.5		9.47	266
12/04/89	05:00	124.5		7.05	264
12/04/89	10:30	124		6.18	289
12/04/89	13:00	124		6.07	292
12/04/89	17:15	124		6.41	293
12/05/89	00:30	122		8.37	259
12/05/89	05:15	123.5		8.11	253
12/05/89	09:30	127		5.55	286
12/05/89	14:15	126.5		6.12	293
12/05/89	19:10	124.5		6.33	288
12/06/89	01:00	121.5		8.17	268
12/06/89	05:00	121.5		8.06	260
12/06/89	09:45	124		6.51	294
12/06/89	10:45	124		6.68	295
12/09/89	23:00	120		8.50	261
12/10/89	04:00	120.5		8.52	258
12/10/89	09:30	121		8.54	259
12/10/89	14:30	119		8.37	260
12/10/89	18:00	121		7.65	264
12/11/89	04:50	122		7.76	265
12/11/89	08:15	126		6.31	282
12/11/89	10:00	125		6.17	287
12/11/89	12:00	124		6.37	292
12/11/89	20:00	125		6.19	283
12/12/89	01:10	121			263
12/12/89	04:45	121		7.83	262
12/14/89	05:00	121		8.66	260
12/14/89	08:00	125		6.59	276
12/14/89	12:00	125.5		6.52	285
01/06/90	01:00	122		8.11	262
01/08/90	19:00	125		5.44	293
01/09/90	19:00	125.5		5.32	295
01/09/90	21:10	125.5		6.16	301
01/09/90	22:45	122		7.60	281
01/10/90	04:45	120.5		7.81	269
01/16/90	14:30	122		6.31	309
01/16/90	16:20	122		6.22	312

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
01/16/90	19:00	122.5		6.03	300
01/16/90	20:54	125.5		5.86	301
01/16/90	23:40	122		8.62	274
01/17/90	09:35	126		5.11	297
01/17/90	11:40	126.5		6.40	287
01/17/90	13:30	126		6.83	284
01/17/90	14:35	126		6.60	280
01/17/90	16:00	126			277
01/17/90	18:45	126.5		5.77	309
01/17/90	20:45	126		5.69	301
01/17/90	23:45	122		8.91	276
01/18/90	04:45	122		8.77	271
01/18/90	22:20	122			286
01/19/90	21:45	120			292
01/19/90	23:30	120		9.73	268
01/20/90	08:30	120		9.27	268
01/20/90	15:45	122		6.90	269
01/20/90	18:15	123		5.23	287
01/20/90	22:15	122		6.53	279
01/21/90	01:45	121		8.70	265
01/21/90	02:15	121		8.66	263
01/21/90	09:30	117.5		8.80	265
01/21/90	11:00	117.5		8.82	266
01/22/90	02:45	118		8.77	266
01/22/90	03:45	118		8.49	266
01/22/90	05:20	121.5		8.39	266
01/22/90	15:00	124		5.41	291
01/22/90	18:30	122.5		5.81	291
01/22/90	20:45	124		5.50	291
01/22/90	23:40	122		6.90	270
01/22/90	23:55	122		6.63	268
01/23/90	04:50	122		7.11	264
01/23/90	13:30	123			
01/23/90	20:26	123.5		5.02	296
01/23/90	23:38	118.5		8.21	268
01/24/90	02:37	119		8.16	267
01/24/90	17:20	122			
01/24/90	20:40	126.5		5.95	300
01/24/90	23:20	122		8.02	278
01/24/90	23:45	121.5		7.96	275
01/25/90	14:00	125.8		5.69	305

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
01/25/90	21:00	126		6.18	292
01/25/90	21:15	126		6.19	292
01/25/90	21:30	126		6.17	293
01/26/90	00:30	121		8.36	268
01/26/90	01:30	121.5		8.38	266
01/26/90	04:30	120		8.47	268
01/26/90	11:30	124		6.27	295
01/26/90	14:56	126.5			286
01/26/90	17:04	123.5			
01/27/90	01:45	120.5		8.39	262
01/27/90	09:02	118		8.61	259
01/27/90	14:00	122		8.59	265
01/27/90	14:45	122		8.82	265
01/28/90	23:45	119		7.98	261
01/29/90	11:55	125		6.59	285
02/01/90	22:30	123.5		8.73	266
02/02/90	00:45	119.5		8.65	258
02/02/90	03:00	121		8.69	255
02/02/90	09:50	125		6.53	283
02/02/90	15:10	118		6.47	283
02/02/90	15:30	119		6.67	283
02/02/90	18:45	123		6.07	282
02/03/90	04:00	118		8.74	254
02/03/90	10:15	124		7.60	263
02/05/90	19:30	123.5		5.85	288
02/05/90	22:45	119.5		8.75	257
02/06/90	07:15	119.5		8.51	258
02/06/90	11:20	122.5		6.49	264
02/06/90	15:30	122.4		6.90	267
02/06/90	18:18	121.5		6.87	267
02/06/90	21:45	119.5		8.02	264
02/06/90	22:15	119.5		7.92	261
02/07/90	02:45	119		8.27	260
02/07/90	03:30	119		8.30	261
02/07/90	04:40	119		8.21	261
02/07/90	11:53	122		6.79	264
02/07/90	16:36	121.6		6.75	267
02/07/90	22:35	118.8		7.58	262
02/08/90	06:45	119.2			
02/08/90	09:00	121		6.13	265
02/08/90	14:45	121		6.98	272

## WET BULB TEMPERATURE MEASUREMENTS

DATE	TIME	MEASURED WET BULB TEMPERATURE (°F)	VOL % O2		HUMIDIFIER INLET GAS TEMPERATURE (°F)
			ESP OUTLET	HUMIDIFIER INLET	
02/08/90	18:40	122		6.65	273
02/08/90	21:00	122.5		5.65	290
02/08/90	22:47	119		8.36	268
02/09/90	02:40	119		8.09	262
02/09/90	10:17	121		7.34	271
02/09/90	20:45	121		7.17	267
02/10/90	00:30	118.5		8.30	258
02/10/90	08:30	117		7.98	253
02/10/90	15:35	117.5		8.89	260
02/10/90	18:55	120			
02/10/90	20:50	120		6.00	288
02/10/90	22:54	117		9.70	265
02/11/90	00:44	117		9.39	257
02/11/90	02:44	117		9.20	258
02/11/90	05:00	116		9.29	259
02/11/90	09:30	117		8.59	259
02/11/90	18:00	116		9.04	262
02/11/90	20:45	116		9.09	262
02/12/90	04:30	117		9.03	259
02/12/90	09:00	125		5.64	287
02/12/90	15:50	122		9.54	262
02/12/90	16:45	119		9.15	259
02/12/90	20:45	124		5.62	286
02/12/90	22:45	119.2			
02/13/90	00:01	119.2		9.29	264
02/13/90	05:30	118.6		8.54	261
02/13/90	16:55	123.5		5.37	296
02/13/90	20:15	125		5.77	287
02/14/90	01:00	118		9.52	251
02/14/90	04:44	115.4		10.16	249
02/14/90	11:40	121		6.13	270
02/14/90	19:00	119		5.76	276
02/14/90	23:45	118		9.00	253
02/15/90	00:40	118.2		9.06	252
02/15/90	22:47	118		8.80	254
02/16/90	00:45	118		8.66	254
02/16/90	03:40	116.6		8.72	257
02/16/90	04:38	116.6		8.75	257
02/16/90	09:12	114.3		8.80	259
02/16/90	11:30	117.3		8.95	260

PROCESS CONDITIONS AND SO<sub>2</sub> REMOVALS

Date	Sorbent Injector Location, Ft from Water Sprays	Start Time	Duration, hrs	Coal Sulfur Content wt %	Baseline SO <sub>2</sub> at 0% Excess Air(1)	Gross Load MW	Average Temperatures, °F		Water/ Flue Gas gal/lb	Stack Gas SO <sub>2</sub> ppm O <sub>2</sub> %	Corrected to 0% Excess Air	SO <sub>2</sub> ppm & SO <sub>2</sub> Removal	Ca/S molar	
							Humidifier Inlet	ESP Inlet						
8/29	-8	1145	1.00	1.65	1693	64.7	146	147	4.0	607	7.30	933	44.9	1.89
8/29	-8	1745	5.17	1.65	1693	61.8	145	144	4.1	482	7.98	781	53.9	1.83
8/30	-8	0225	3.00	1.65	1693	45.7	145	148	3.7	501	9.35	907	46.4	1.77
8/30	-8	0645	2.00	1.65	1693	64.7	145	142	4.0	496	7.67	782	53.8	1.83
9/7	0	0000	2.33	1.76	1780	42.9	287	151	3.8	733	7.91	1180	33.7	1.81
9/7	0	0230	4.33	1.76	1780	43.3	286	148	3.0	686	8.05	1115	37.4	2.02
9/8	0	0150	4.17	1.66	1677	43.1	287	151	3.0	644	9.03	1134	32.4	1.86
9/8	0	0810	1.17	1.66	1677	61.7	297	150	4.1	683	8.45	1146	31.6	1.97
9/11	0	1200	2.00	1.67	1715	65.8	291	145	3.9	636	8.09	1038	39.5	1.73
9/11	0	1400	3.00	1.67	1715	66.1	289	146	3.8	681	8.01	1106	35.5	1.74
9/11	0	2010	2.00	1.67	1715	66.6	291	145	4.0	674	7.91	1085	36.7	1.76
9/12	0	0140	3.67	1.66	1707	43.8	272	147	3.4	691	8.90	1204	29.5	1.68
9/12	0	1030	2.33	1.66	1707	35.3	272	146	3.8	521	10.09	1006	41.1	1.56
9/12	0	1915	1.50	1.66	1707	66.4	289	145	3.8	614	8.01	995	41.7	1.79
9/12	0	2115	2.75	1.66	1707	44.0	285	148	3.5	575	9.01	1012	40.7	1.66
9/13	0	0006	1.83	1.45	1488	44.9	281	148	3.4	622	9.10	1101	26.0	1.95
9/13	0	0346	3.00	1.45	1488	44.3	278	145	3.3	588	9.06	1037	30.3	1.92
9/13	0	1526	3.83	1.45	1488	45.7	275	152	3.1	625	8.84	1083	27.2	1.93
9/18	0	2300	1.00	1.76	1805	46.1	287	145	3.6	460	9.59	850	52.9	2.51
9/19	0	0000	6.00	1.71	1754	46.2	282	145	3.6	508	9.44	926	47.2	2.59
9/19	0	1636	1.28	1.71	1754	59.7	291	153	3.0	611	8.97	1070	39.0	1.36
9/19	0	1815	3.12	1.71	1754	62.4	291	145	4.0	589	9.04	1037	40.9	1.38
9/19	0	2206	1.10	1.71	1754	44.1	288	145	3.5	661	9.24	1185	32.5	1.29
9/28	0	1158	1.00	1.63	1671	63.7	291	144	3.9	448	8.75	772	53.8	3.02
9/28	0	1420	2.38	1.63	1671	58.5	290	145	3.8	475	9.05	837	49.9	2.96

All gas concentrations are dry basis.

(1) Based on average coal sulfur content.

Early process runs data for Figure 8.

## Pulverized Coal Samples – Commercial Testing &amp; Engineering Analyses

----- AS RECEIVED -----

DATE	SAMPLE NUMBER	VOLATILE MATTER	FIXED CARBON	H <sub>2</sub> O	ASH	S	HHV	C	H	N	O	COMMENT
880830	LIMB765	32.25	49.81	3.74	14.20	1.99	11976	67.93	4.25	1.33	6.56	Low S
880831	LIMB766	33.21	49.13	2.96	14.70	2.17	11983	66.67	4.53	1.31	7.66	Low S
880901	LIMB767	32.87	49.77	2.88	14.48	2.19	12013	68.52	4.62	1.30	6.01	Low S
880902	LIMB768	33.14	50.50	2.99	13.37	1.96	12241	68.51	4.52	1.37	7.28	Low S
880927	LIMB787	36.61	50.36	3.67	9.36	2.26	12460	70.10	4.74	1.41	8.46	Low S
881020	LIMB830	35.82	49.78	4.91	9.49	2.21	12528	69.43	4.64	1.39	7.93	Low S residual
881117	LIMB932	34.83	49.34	4.05	11.78	2.26	12366	68.81	4.65	1.51	6.94	Low S
881122	LIMB939	31.48	48.15	8.52	11.85	1.58	11761	65.34	4.30	1.37	7.04	Low S
881123	LIMB943	32.71	49.53	6.54	11.22	1.63	11979	66.77	4.57	1.40	7.87	Low S
881128	LIMB947	33.61	50.45	4.57	11.37	1.67	12214	68.37	4.67	1.40	7.95	Low S
881129	LIMB951	30.61	52.90	4.35	12.14	1.71	12255	67.92	4.62	1.41	7.85	Low S
881130	LIMB955	30.31	53.70	3.88	12.11	1.57	12259	69.12	4.49	1.47	7.36	Low S
890601	LIMB1606	34.72	49.48	4.12	11.68	2.59	12345	68.76	4.64	1.45	6.76	High S
890602	LIMB1614	34.41	48.70	4.42	12.47	2.59	12179	67.50	4.48	1.44	7.10	High S
890605	LIMB1626	34.06	49.45	4.49	12.00	2.59	11966	67.12	4.44	1.43	7.93	High S
890606	LIMB1632	36.07	49.12	4.35	10.46	2.87	12473	69.09	4.62	1.51	7.10	High S
890607	LIMB1642	35.98	48.81	4.30	10.91	3.16	12480	68.97	4.71	1.32	6.63	High S
890608	LIMB1650	37.72	49.50	3.64	9.14	2.83	12851	70.54	4.80	1.48	7.57	High S
890609	LIMB1655	37.24	50.35	3.46	8.95	2.54	12793	70.97	4.78	1.51	7.79	High S, unit off-line PM
890719	LIMB1668	32.76	52.15	4.12	10.97	1.35	12145	68.57	4.77	1.42	8.80	Low S
890720	LIMB1672	33.06	50.42	4.34	12.18	1.44	11812	67.28	4.39	1.39	8.98	Low S
890721	LIMB1676	33.34	50.05	4.69	11.92	1.27	11918	67.27	4.37	1.41	9.07	Low S
890728	LIMB1685	32.06	51.99	4.40	11.55	1.53	12115	68.13	4.68	1.37	8.34	Low S
890801	LIMB1691	32.81	52.31	4.08	10.80	1.57	12311	69.28	4.78	1.39	8.10	Low S
890802	LIMB1695	32.21	52.49	3.69	11.61	1.71	12280	68.65	4.78	1.40	8.16	Low S
890803	LIMB1699	33.27	52.54	3.52	10.67	1.49	12485	69.90	4.93	1.38	8.11	Low S
890804	LIMB1702	32.69	53.09	3.28	10.94	1.49	12518	70.07	4.78	1.43	8.01	Low S
890807	LIMB1708	34.90	49.41	3.60	12.09	1.52	12210	68.41	4.43	1.39	8.56	Low S
890808	LIMB1715	32.22	52.61	3.49	11.68	1.43	12301	69.49	4.56	1.40	7.95	Low S; tube leak, unit down
890813	LIMB1722	31.80	52.73	3.66	11.81	1.32	12200	69.26	4.49	1.45	8.01	Low S
890815	LIMB1729	33.40	51.72	3.18	11.70	1.55	12355	69.70	4.50	1.49	7.88	Low S
890816	LIMB1735	34.10	51.23	3.40	11.27	1.47	12401	69.54	4.67	1.45	8.20	Low S
890817	LIMB1742	33.01	51.22	3.26	12.51	1.59	12395	68.67	4.77	1.40	7.80	Low S
890818	LIMB1747	31.92	52.05	3.40	12.63	1.54	12232	68.54	4.54	1.54	7.81	Low S
890821	LIMB1752	31.52	51.72	3.51	13.25	1.56	12056	67.55	4.54	1.53	8.06	Low S
890822	LIMB1757	32.30	51.14	3.54	13.02	1.50	12155	68.03	4.59	1.46	7.86	Low S
890823	LIMB1762	32.21	51.84	3.38	12.57	1.56	12092	68.42	4.45	1.38	8.24	Low S
890824	LIMB1767	32.29	51.53	3.65	12.53	1.51	12146	68.10	4.48	1.44	8.29	Low S
890828	LIMB1774	31.95	51.63	3.43	12.99	1.50	12090	67.95	4.52	1.44	8.17	Low S
890829	LIMB1784	32.76	51.54	3.38	12.32	1.75	12276	68.11	4.78	1.44	8.22	Low S
890830	LIMB1789	32.39	52.10	3.32	12.19	1.43	12298	68.81	4.73	1.51	8.01	Low S
890901	LIMB1795	32.37	51.26	3.48	12.89	1.57	12124	67.97	4.62	1.44	8.03	Low S
890905	LIMB1800	32.12	51.69	3.50	12.69	1.40	12118	68.06	4.62	1.39	8.34	Low S
890906	LIMB1807	32.54	51.08	3.94	12.44	1.58	12122	67.84	4.65	1.42	8.13	Low S
890907	LIMB1812	33.33	51.81	3.13	11.73	1.58	12421	69.43	4.75	1.41	7.97	Low S
890908	LIMB1819	32.30	52.03	3.25	12.42	1.51	12275	68.98	4.48	1.47	7.89	Low S
890911	LIMB1828	32.40	50.70	3.70	13.20	1.51	11990	67.82	4.37	1.48	7.92	Low S
890912	LIMB1839	32.88	50.18	3.70	13.24	1.26	11992	67.38	4.36	1.38	8.68	Low S
890913	LIMB1848	32.53	51.41	3.61	12.45	1.26	12194	68.80	4.37	1.37	8.14	Low S
890914	LIMB1858	31.55	52.47	3.65	12.33	1.49	12184	68.50	4.45	1.49	8.09	Low S
890915	LIMB1864	31.89	52.27	3.53	12.31	1.57	12240	68.73	4.48	1.56	7.82	Low S
890918	LIMB1871	32.59	51.19	4.13	12.09	1.52	14400	67.81	4.36	1.47	8.62	Low S
890919	LIMB1878	32.58	50.46	4.12	12.84	1.53	12113	68.06	4.39	1.54	7.52	Low S
890920	LIMB1886	31.89	52.00	3.71	12.40	1.48	12221	68.21	4.55	1.46	8.19	Low S
890921	LIMB1891	32.36	51.94	3.63	12.07	1.44	12269	68.52	4.71	1.49	8.14	Low S
890922	LIMB1897	32.86	52.41	3.27	11.46	1.40	12553	69.94	4.80	1.55	7.58	Low S

Pulverized Coal Samples – Commercial Testing & Engineering Analyses

----- AS RECEIVED -----												
DATE	SAMPLE NUMBER	VOLATILE MATTER	FIXED CARBON	H <sub>2</sub> O	ASH	S	HHV	C	H	N	O	COMMENT
890925	LIMB1905	35.62	49.29	3.69	11.40	1.51	12455	69.04	4.75	1.51	8.10	Low S
890926	LIMB1914	32.97	52.58	3.31	11.14	1.52	12465	69.97	4.82	1.50	7.74	Low S
890927	LIMB1925	32.47	52.66	3.35	11.52	1.36	12481	69.80	4.94	1.49	7.54	Low S
890928	LIMB1934	32.73	52.59	3.66	11.02	1.43	12481	69.85	4.79	1.54	7.71	Low S
890929	LIMB1938	32.55	52.39	3.58	11.48	1.45	12456	69.69	4.68	1.50	7.62	Low S
891002	LIMB1946	32.88	51.49	3.76	11.87	1.65	12322	69.13	4.83	1.52	7.44	Start of High S
891003	LIMB1960	33.66	51.67	3.64	11.03	2.15	12414	69.00	4.70	1.40	8.08	High S
891004	LIMB1973	33.93	50.58	4.29	11.20	2.26	12175	67.83	4.76	1.49	8.17	High S
891005	LIMB1985	34.22	50.66	3.84	11.28	2.38	12246	68.28	4.68	1.48	8.06	High S
891006	LIMB2001	33.44	49.40	3.42	13.74	2.62	11944	66.45	4.66	1.46	7.65	High S
891009	LIMB2007	34.53	50.42	3.40	11.65	2.50	12309	68.14	4.69	1.44	8.18	High S
891010	LIMB2019	34.25	50.49	3.60	11.66	2.46	12331	67.91	4.73	1.44	8.20	High S
891011	LIMB2030	33.86	50.06	4.71	11.37	2.54	12025	67.18	4.57	1.35	8.28	High S
891012	LIMB2041	34.83	50.41	4.11	10.65	2.58	12185	68.15	4.62	1.38	8.51	High S
891013	LIMB2050	35.34	49.66	3.69	11.31	2.58	12242	67.99	4.66	1.46	8.31	High S
891016	LIMB2058	42.69	42.35	3.97	10.99	2.51	12276	68.21	4.67	1.44	8.21	High S
891017	LIMB2065	35.34	50.55	3.80	10.31	2.49	12467	69.13	4.78	1.41	8.08	High S
891018	LIMB2076	36.40	49.49	3.39	10.72	2.81	12591	69.36	4.80	1.37	7.55	High S
891019	LIMB2086	36.79	50.39	2.77	10.05	2.71	12797	70.54	4.86	1.31	7.76	High S
891020	LIMB2091	34.99	50.73	3.29	10.99	2.38	12628	69.52	4.72	1.30	7.80	High S
891023	LIMB2095	35.22	51.31	3.24	10.23	2.16	12682	70.00	4.90	1.35	8.12	Low S loaded to bunker
891025	LIMB2102	33.61	52.40	3.74	10.25	1.84	12556	70.09	4.90	1.43	7.75	Low S
891026	LIMB2114	33.68	52.45	3.53	10.34	1.65	12501	70.33	4.93	1.43	7.79	Low S
891027	LIMB2120	33.70	52.59	3.71	10.00	1.72	12625	70.84	4.77	1.44	7.52	Low S
891030	LIMB2125	33.50	52.68	3.43	10.39	1.57	12652	70.67	4.85	1.44	7.85	Low S
891031	LIMB2133	33.64	52.68	3.46	10.22	1.53	12571	70.67	4.70	1.54	7.88	Low S
891101	LIMB2142	32.90	52.28	4.04	10.78	1.38	12353	69.67	4.75	1.46	7.92	Low S
891102	LIMB2151	32.76	52.31	3.87	11.06	1.42	12402	69.93	4.62	1.45	7.65	Low S
891106	LIMB2156	32.73	52.45	3.91	10.91	1.38	12419	69.84	4.64	1.50	7.82	Low S; High S loaded
891111	LIMB2168	33.99	51.01	3.57	11.43	2.27	12316	68.79	4.71	1.38	7.85	High S
891112	LIMB2166	35.46	50.21	3.48	10.85	2.28	12498	69.53	4.80	1.44	7.62	High S
891113	LIMB2177	34.78	52.42	3.38	9.42	1.79	12626	71.52	4.98	1.54	7.37	High S
891114	LIMB2185	35.31	50.49	3.03	11.17	2.61	12515	69.81	4.91	1.44	7.03	High S
891115	LIMB2191	40.74	46.94	2.66	9.66	2.76	12781	70.69	4.97	1.42	7.84	High S
891116	LIMB2197	33.12	52.53	3.90	10.45	1.61	12548	70.01	4.82	1.45	7.76	Low S loaded to bunker
891117	LIMB2202	32.36	52.78	4.12	10.74	1.37	12532	69.92	4.68	1.40	7.77	Low S
891120	LIMB2207	34.71	50.68	3.89	10.72	1.37	12499	70.52	4.76	1.43	7.31	Low S
891121	LIMB2212	33.06	52.05	3.70	11.19	1.23	12443	70.10	4.76	1.43	7.59	Low S
891122	LIMB2219	33.46	51.42	4.07	11.05	1.15	12302	69.46	4.68	1.42	8.17	Low S
891127	LIMB2222	32.93	52.34	3.44	11.29	1.36	12460	70.04	4.64	1.43	7.80	Low S
891128	LIMB2233	34.30	50.59	3.38	11.73	1.61	12485	69.74	4.82	1.43	7.29	Low S; High S coal loaded
891129	LIMB2240	36.38	49.56	3.85	10.21	2.45	12554	69.44	4.91	1.36	7.78	High S
891130	LIMB2248	36.31	49.54	3.69	10.46	2.74	12404	69.16	4.79	1.43	7.73	High S
891201	LIMB2255	39.69	46.21	3.52	10.58	2.77	12485	69.31	4.89	1.39	7.54	High S
891202	LIMB2261	36.70	49.36	3.36	10.58	2.91	12519	69.26	4.93	1.39	7.57	High S
891203	LIMB2268	38.22	49.08	3.33	9.37	2.86	12774	70.51	5.05	1.37	7.51	High S
891204	LIMB2279	36.43	49.96	3.71	9.90	2.58	12665	70.01	4.96	1.43	7.41	High S
891205	LIMB2291	40.48	44.86	4.01	10.65	2.79	12516	69.10	4.81	1.42	7.22	High S
891206	LIMB2300	36.81	50.01	3.36	9.82	2.60	12671	70.37	4.81	1.51	7.53	High S; Low S loaded
891207	LIMB2305	34.44	51.06	3.86	10.64	1.69	12530	70.23	4.63	1.52	7.43	Low S
891208	LIMB2309	36.00	50.56	3.49	9.95	1.71	12813	70.71	4.74	1.44	7.96	Low S
891210	LIMB2316	38.79	47.28	3.68	10.25	2.80	12533	69.28	4.75	1.34	7.90	High S loaded to bunker
891211	LIMB2326	38.99	48.42	3.67	8.92	2.53	12747	70.90	4.78	1.41	7.79	High S
891212	LIMB2333	38.55	48.00	3.58	9.87	3.00	12610	69.91	4.79	1.45	7.40	High S
891213	LIMB2339	39.17	48.18	3.33	9.32	2.93	12780	70.49	4.80	1.60	7.53	High S
891214	LIMB2348	38.03	47.68	3.50	10.79	2.81	12511	69.34	4.69	1.51	7.36	High S

Pulverized Coal Samples – Commercial Testing & Engineering Analyses

----- AS RECEIVED -----

DATE	SAMPLE NUMBER	VOLATILE MATTER	FIXED CARBON	H <sub>2</sub> O	ASH	S	HHV	C	H	N	O	COMMENT
891219	LIMB2353	36.94	48.20	3.96	11.50	2.81	12379	69.02	4.69	1.36	7.26	High S
891221	LIMB2357	38.37	46.98	3.81	11.04	2.52	12483	69.08	4.90	1.37	7.48	High S
900102	LIMB2365	34.34	50.73	4.30	10.63	1.33	12398	69.66	4.66	1.43	7.99	Low S
900103	LIMB2370	35.15	51.14	3.96	9.75	1.23	12645	70.67	4.61	1.47	8.31	Low S
900104	LIMB2375	34.04	51.32	3.56	11.08	1.35	12605	70.20	4.62	1.48	7.71	Low S
900108	LIMB2382	34.24	52.49	3.54	9.73	1.49	12751	71.24	4.77	1.51	7.72	Low S
900109	LIMB2393	32.70	51.91	3.10	12.29	1.36	12441	69.71	4.64	1.48	7.42	Low S
900110	LIMB2401	32.43	52.87	2.86	11.84	1.33	12621	70.70	4.67	1.46	6.99	Low S
900111	LIMB2411	34.26	52.43	2.81	10.50	1.26	11958	67.64	4.98	1.32	*****	Low S
900112	LIMB2417	34.24	51.59	2.80	11.37	1.29	12706	70.92	4.69	1.39	7.38	Low S
900115	LIMB2422	32.59	53.23	2.96	11.22	1.56	12693	71.08	4.72	1.44	6.87	Low S
900116	LIMB2430	33.40	53.02	3.09	10.49	1.31	12780	71.61	4.77	1.42	7.16	Low S
900117	LIMB2440	32.67	52.77	4.03	10.53	1.31	12542	70.67	4.63	1.54	7.17	Low S
900118	LIMB2449	32.13	52.78	4.40	10.69	1.33	12417	70.07	4.60	1.53	7.27	Low S
900119	LIMB2458	33.12	53.09	3.91	9.88	1.23	12648	71.53	4.58	1.57	7.17	Low S
900120	LIMB2464	34.20	52.75	3.62	9.43	1.20	12801	71.58	4.57	1.56	7.92	Low S
900122	LIMB2477	33.67	52.10	3.86	10.37	1.36	12640	70.57	4.51	1.52	7.65	Low S
900123	LIMB2487	33.73	52.75	4.07	9.45	1.37	12514	69.96	4.63	1.65	8.73	Low S
900124	LIMB2498	34.72	49.26	3.61	12.41	1.66	12419	68.95	4.43	1.58	7.22	Low S
900125	LIMB2510	34.32	49.60	3.33	12.75	2.46	12262	68.08	4.50	1.54	7.19	High S loaded to bunker AM
900126	LIMB2520	36.72	47.53	3.17	12.58	2.56	12367	68.30	4.54	1.55	7.12	High S
900129	LIMB2526	33.99	49.94	3.69	12.38	2.64	12291	67.82	4.58	1.32	7.41	High S
900201	LIMB2531	34.04	49.80	3.78	12.38	2.79	12147	67.43	4.48	1.29	7.69	High S
900202	LIMB2539	34.10	51.05	3.78	11.07	2.30	12454	68.97	4.52	1.34	7.87	High S
900203	LIMB2547	32.52	52.85	4.22	10.41	1.47	12541	70.01	4.48	1.40	7.87	Low S
900205	LIMB2554	32.84	52.34	4.29	10.53	1.28	12519	69.85	4.47	1.45	8.00	Low S
900206	LIMB2564	32.93	52.72	4.08	10.27	1.34	12568	70.15	4.47	1.51	8.00	Low S
900207	LIMB2573	33.53	52.37	4.03	10.07	1.42	12623	70.59	4.59	1.53	7.61	Low S
900208	LIMB2583	33.49	52.36	3.87	10.28	1.41	12613	70.66	4.44	1.50	7.72	Low S
900209	LIMB2592	35.06	51.75	3.59	9.60	1.44	12734	71.55	4.68	1.49	7.53	Low S
900210	LIMB2597	34.32	51.89	3.82	9.97	1.44	12646	70.93	4.49	1.46	7.78	Low S
900211	LIMB2602	33.57	52.20	3.88	10.35	1.25	12548	70.32	4.39	1.44	8.25	Low S
900212	LIMB2611	33.67	52.54	3.96	9.83	1.18	12685	70.98	4.86	1.49	7.58	Low S
900213	LIMB2619	33.11	52.11	3.78	11.00	1.35	12545	70.05	4.73	1.11	7.84	Low S
900214	LIMB2626	32.97	52.93	3.95	10.15	1.27	12648	70.85	4.72	1.53	7.37	Low S
900215	LIMB2634	32.58	52.48	3.94	11.00	1.24	12466	70.05	4.72	1.52	7.36	Low S
900216	LIMB2641	34.52	52.92	3.50	9.06	1.31	12889	71.56	4.72	1.59	8.10	Low S, END OF COOLSIDE

**APPENDIX B. CALIBRATION DATA**

**TABLE B-1.  
HUMIDIFIER FLUE GAS VELOCITIES (PITOT TUBE MEASUREMENTS)  
EDGEWATER STATION -- COOLSIDE PROCESS TESTS**

Test 1								
Description	Case 1 - High Velocity Run Without Atomizing Air -- Humidifier Inlet							
Date	7/20/89							
Time	10:39-12:48 hrs.							
Avg. Temp., °F	296							
Distance, Ft. From North Wall Above Floor	1.3	3.3	5.3	7.3	9.3	11.3	13.3	
	Flue Gas Velocity, Ft./Sec.							
13.54	23.0	24.5	23.3	20.4	22.2	25.8	33.0	
11.46	20.5	25.0	22.8	21.5	22.4	22.1	34.5	
9.38	19.5	24.1	24.7	23.5	26.5	29.3	35.2	
7.29	20.2	26.7	25.2	23.8	22.8	22.8	34.5	
5.21	19.8	23.0	27.0	26.4	26.7	28.2	35.8	
3.13	22.2	24.8	22.8	31.1	20.8	27.5	32.8	
1.04	18.9	22.8	20.8	31.6	23.1	27.1	33.1	

Test 2								
Description	Case 1 - High Velocity Run Without Atomizing Air -- Humidifier Center							
Date	7/20/89							
Time	13:12-14:35 hrs.							
Avg. Temp., °F	288							
Distance, Ft. From North Wall Above Floor	0.5	3.3	5.3	7.3	9.3	11.3	14.1	
	Flue Gas Velocity, Ft./Sec.							
13.54	21.6	21.0	23.6	22.3	22.8	25.0	32.3	
11.46	20.0	25.5	23.4	23.3	23.9	26.6	32.0	
9.38	23.8	26.1	24.6	25.0	27.2	30.8	33.7	
7.29	27.6	26.6	23.9	26.3	29.8	33.7	30.6	
5.21	23.4	26.3	27.0	26.5	29.2	32.4	33.2	
3.13	20.9	25.9	25.6	26.1	27.6	28.4	33.1	
1.04	17.4	19.5	19.8	24.0	22.3	25.7	27.5	

(continued)

**APPENDIX B. CALIBRATION DATA**

**TABLE B-1. (continued)**

Test 3							
Description	Case 1 - High Velocity Run Without Atomizing Air -- Humidifier Outlet						
Date	7/20/89						
Time	14:50-15:30 hrs						
Avg. Temp., °F	285						
Distance, Ft. From North Wall Above Floor	1.3	3.3	5.3	7.3	9.3	11.3	13.3
	Flue Gas Velocity, Ft./Sec.						
13.54	21.4			21.2			26.3
11.46	26.9			28.5			32
9.38	29.3			31.8			35.1
7.29	31.8			34.9			35.4
5.21	28.8			33.0			35.8
3.13	23.7			29.0			33.5
1.04	14.0			16.3			17.4

Test 4							
Description	Case 2 - Low Velocity Run Without Atomizing Air -- Humidifier Inlet						
Date	7/21/89						
Time	10:11-11:31 hrs						
Avg. Temp., °F	292						
Distance, Ft. From North Wall Above Floor	1.3	3.3	5.3	7.3	9.3	11.3	13.3
	Flue Gas Velocity, Ft./Sec.						
13.54	14.8	17.8	18.2	15.3	14.4	18.4	21.5
11.46	15.0	18.4	15.6	15.0	16.2	17.2	24.3
9.38	14.9	15.9	17.8	16.7	19.1	19.9	24.3
7.29	14.5	18.5	18.4	17.7	17.5	17.6	23.8
5.21	15.8	17.5	18.4	18.2	19.1	19.8	24.6
3.13	13.7	18.0	15.5	20.8	16.0	19.4	23.6
1.04	15.6	18.0	15.6	20.2	15.9	19.7	22.7

(continued)

**APPENDIX B. CALIBRATION DATA**

**TABLE B-1. (continued)**

Test 5								
Description	Case 2 - Low Velocity Run Without Atomizing Air -- Humidifier Center							
Date	7/21/89							
Time	11:40-12:50 hrs							
Avg. Temp., °F	287							
Distance, Ft. From North Wall Above Floor	0.5	3.3	5.3	7.3	9.3	11.3	14.1	
	Flue Gas Velocity, Ft./Sec.							
13.54	16.3	16.9	17.7	16.4	15.9	17.8	23.1	
11.46	16.1	20.2	19.6	17.4	18.9	19.5	23.9	
9.38	18.0	20.0	19.1	18.5	20.3	21.2	23.7	
7.29	22.1	18.7	17.4	19.7	22.6	25.4	22.4	
5.21	18.1	20.1	19.7	19.2	21.3	23.5	23.6	
3.13	15.1	19.9	19.2	19.5	19.8	20.7	24.1	
1.04	13.2	14.3	15.1	17.9	16.9	18.2	14.0	

Test 6								
Description	Case 3 - Medium Velocity Run With Atomizing Air -- Humidifier Inlet							
Date	7/21/89							
Time	17:05-18:07 hrs							
Avg. Temp., °F	287							
Distance, Ft. From North Wall Above Floor	1.3	3.3	5.3	7.3	9.3	11.3	13.3	
	Flue Gas Velocity, Ft./Sec.							
13.54	16.0	18.0	20.4	17.2	18.9	22.4	27.6	
11.46	15.7	16.8	20.9	20.3	20.6	20.5	30.2	
9.38	17.2	16.8	21.9	22.0	25.8	25.0	31.2	
7.29	19.1	24.2	24.3	20.8	23.7	24.7	30.8	
5.21	13.6	19.5	22.5	23.9	25.0	26.4	29.9	
3.13	16.3	20.2	19.9	23.9	20.8	22.3	29.1	
1.04	8.4	15.8	17.3	21.4	19.9	23.6	27.0	

(continued)

**APPENDIX B. CALIBRATION DATA**

**TABLE B-1. (continued)**

<b>Test 7</b>								
<b>Description</b>	<b>Case 3 - Medium Velocity Run With Atomizing Air -- Humidifier Center</b>							
<b>Date</b>	7/21/89							
<b>Time</b>	15:37-16:40 hrs							
<b>Avg. Temp., °F</b>	280							
<b>Distance, Ft. From North Wall Above Floor</b>	0.5	3.3	5.3	7.3	9.3	11.3	14.1	
	<b>Flue Gas Velocity, Ft./Sec.</b>							
13.54	-4.1	15.6	23.6	26.2	26.6	21.8	-2.8	
11.46	6.1	20.4	39.2	44.9	47.4	31.7	6.1	
9.38	13.7	25.4	38.9	44.4	47.1	32.9	10.6	
7.29	23.5	33.8	42.6	47.1	49.6	32.1	12.1	
5.21	23.0	35.5	43.3	48.2	49.1	34.7	14.0	
3.13	10.3	16.1	26.0	30.3	31.1	18.5	-6.3	
1.04	3.0	-4.2	6.3	8.1	9.1	3.8	-9.1	

APPENDIX B. CALIBRATION DATA

TABLE B-2  
 RECYCLE SOLIDS CONVEYING TEST  
 SOLIDS FEED TO DISTRIBUTION BOTTLE - 5/31/89

Time	16:02	16:28	16:50	18:46	19:10	19:33	19:45	19:58	20:15	20:33	21:35
<b>Air Compressor</b>											
Pres., psi	56	56	63	63	63	63	63	63	63	63	82
Temp., °F	181	180	182	185	185	185	185	184	184	184	184
Air Flow, cfm*	715	715	715	715	715	715	715	715	715	715	715
<b>Rotary Feeder</b>											
Speed, rpm	0.0	1.2	2.2	3.2	6.0	9.5	12.5	15.8	17.0	12.5	9.5
<b>Solids Pickup Pres., psi</b>											
Low	0.00	0.77	1.50	2.55	4.36	5.52	6.50	7.53	8.27	6.57	5.50
High	0.00	1.50	2.20	2.55	4.36	5.69	6.80	7.84	8.60	6.80	5.88
Average	0.00	1.14	1.85	2.55	4.36	5.61	6.65	7.69	8.44	6.69	5.69
<b>Distribution Bottle Pres., psi</b>											
Low	0.00	0.20	0.50	1.10	1.40	1.80	2.10	2.40	2.60	2.00	2.00
High	0.00	0.40	0.60	1.50	2.00	2.50	3.30	4.00	4.50	3.20	2.50
Average	0.00	0.30	0.55	1.30	1.70	2.15	2.70	3.20	3.55	2.60	2.25
Transfer Line DP, psi	0.00	0.84	1.30	1.25	2.66	3.46	3.95	4.49	4.89	4.09	3.44

\* Actual flow based on compressor specifications and pitot tube measured check of compressor capacity.





- 7/17 process equipment was completed.
- 7/18 Floor dust blower pipes were installed in the humidifier. The purpose of these pipes was to help convey out of the humidifier the solids which drop from the flue gases on to the humidifier floor.
- 7/20 Began preliminary Coolside process tests to checkout the unit operation. The first tests were to measure the flue gas velocity profiles into, in, and out of the humidifier. Completed velocity measurements at a boiler load of 75 MW.
- 7/21 Measured duct flue gas velocities at reduced boiler load and determined the effect of the water spray atomizer high pressure air flow on the flue gas velocity profiles in the humidifier.
- 7/24 Checked the air flow rates to the individual floor dust blower pipes using the atomizing air flow supply meter. The measured rates agreed closely with calculations.
- The boiler was taken off line for an instrument check.
- 7/25 During the afternoon hours, the Coolside process equipment was brought on-line. Because of hydrated lime leaks, the 'B' lime feeder was taken out of service and the 'A' feeder placed in service.
- The Coolside process equipment was shut down when the hydrated lime day silo overfilled causing lime dust to blow out of the silo baghouse access door. The overfilling was due to a problem with the automatic fill cycle electronics. The Coolside system remained shut down to allow cleanup of the spill.
- 7/27 As an added precaution against future overfill occurrences, an additional fill-stop interlock using the day silo weight was installed.

7/28 Several of the water spray lance array high-pressure air hoses were found to be leaking where the hoses attach to the metal lances. The rubber had hardened and cracked. Repairs were made by cutting off the hardened rubber ends and then re-banding the hoses to the lances.

Attempted a Coolside process test at Ca/S = 1.0, Na/Ca = 0.0, and 20°F approach. Obtaining stable operating conditions was difficult due to the instrument tuning of the newly installed boiler electronic control system. Coolside process tests were curtailed when one of the rotary air seal valves ahead of the hydrated lime gravimetric weigh feeders failed to operate. The problem was due to a bad limit switch which was repaired on 7/31.

7/29 Boiler off line for weekend. Electrical generation not needed.  
to  
7/30

7/31 The boiler was back on-line burning low-sulfur coal during the afternoon. Because of the boiler control tuning activities, Coolside process operations were limited to short tests. Tests were conducted at Ca/S = 2.0, Na/Ca = 0.0, and 20°F approach using the first set of solids injector lances. The first set of solids injectors was designed to provide the widest solids dispersion across the duct ahead of the water atomizer lances.

8/1 Tests were conducted to determine if the Coolside process equipment was affecting the boiler controls. The control problems were found not to be related to the Coolside operations.

At 17:00 hrs, the boiler tripped and the system was shut down.

8/2 The hydrated lime conveying air compressor regulating solenoid valve malfunctioned causing the compressor discharge pressure to cycle. The Coolside equipment was shut down to allow repair.

8/3 The hydrated lime compressor output was controlled by manually adjusting the position of the suction throttle butterfly valve until the problem with the solenoid controls could be resolved.

Conducted a Coolside process test using the first or wide dispersion set of solids injectors. Conditions were: Ca/S = 2.5, Na/Ca = 0.0, and 20°F approach.

Checked accuracy of Day silo weigh scale by placing known weight (several weighed workers) on silo.

Conducted special recycle solids feed rate tests by transferring solids from the ash silo into the hydrated lime day silo. Used the weight change of the day silo to calibrate feeder RPMs with transfer rate.

8/4 Disposed of the ash solids in the hydrated lime day silo by feeding the solids into the humidifier through the hydrated lime injection system.

Removed the first set of solids injectors from the humidifier and installed the alternate injector set. This second injector set was designed to confine the solids dispersion to the projected area of the water spray atomizer array.

Inspection of the humidifier showed that loose piles of hydrated lime dust had laid down on the water spray lances and reduced the gas flow area between the lances. Some of this material had fallen off the lances and accumulated on the floor just in front and just behind the water sprays. Floor dust blowers in these areas were overwhelmed by the amount of solids. The floor blowers downstream of the atomizer array had kept the floor free of solids.

8/5 Boiler down for weekend.

to

8/6

8/7 The hydrated lime feed line from storage to the day silo was reconnected following the special recycle solids feeder calibration tests.

Began testing of the second set of solids injectors at: Ca/S = 2.0, Na/Ca = 0.0, and 20°F approach.

At 17:00 hrs, the hydrated lime feeder shut down because of a false instrument signal which indicated a low hydrated lime conveying air flow. Again inspected the humidifier and found large dust buildups on the atomizing water spray lances.

8/8 Reestablished Coolside process test conditions at Ca/S = 1.0, Na/Ca = 0.0, and 20°F approach.

Relative accuracy tests were initiated to establish the accuracy of the gas analyzers.

Boiler was shut down due to a steam tube leak.

8/9 While the boiler was down for repairs, inspected the humidifier and to again found large dust accumulations on the water spray lances.

8/13 Removed the distribution disks at the solids injector pipe outlets to determine if the impaction of the hydrated lime on these disks was causing the lime to agglomerate. Particle agglomeration could have been responsible for the dust laydown on the water sprays. Also, the hydrated lime conveying hoses were grounded to eliminate static electric effects. These equipment changes later proved to be ineffective.

8/14 Conducted one Coolside process test at Ca/S = 2.0 and began second test at Ca/S = 1.0. Plant requested a cessation of testing until repairs were completed on the waste water neutralization system.

8/15 At 14:00 hrs, start-up of the Coolside process equipment was initiated. The humidifier shut down at 17:58 hrs when atomizing air flow was

lost for no apparent reason. Because of the late start-up and due to boiler control problems, testing was halted until the following day.

One relative accuracy test was completed.

8/16 Reestablished Coolside test conditions to allow continuation of the relative accuracy gas analyzer tests, test the performance of the second set of solids injectors and to empty out the inventory of hydrated lime 'H' which was left over from the preceding LIMB test program.

8/17 Continued relative accuracy gas analyzer tests.

The ash conveying system dust collector baghouse was opened for inspection because of continuing problems with high bag differential pressures. Several bags were found which appeared to have gotten wet. The ash conveying system was then checked out to insure that highly humidified flue gases were not being drawn into the vacuum system at the end of each ESP hopper dump cycle when the hoppers are empty and there is not a solids seal against gas flow.

The high pressure atomizing air compressor unexpectedly unloaded at 20:31 hrs causing the humidification system to be shut down.

8/18 Coolside process operations were continued to empty out the hydrated lime 'H' inventory. The inventory was exhausted at 17:17 hrs.

The results of the relative accuracy tests confirmed the stack gas analyzer accuracy. The humidifier inlet analyzer was not operating properly and needed repair. Parts were ordered.

8/19 The Coolside process was shut down to allow the installation of a dust  
to blower system for automatic removal of the hydrated lime accumulations  
8/28 on the water spray lances and for boiler steam tube leak repairs. Two additional humidifier floor dust blowers were also installed, one 3' upstream and one 7.5' downstream of the atomizer array.

- 8/28 System checkout and operation using new hydrated lime 'A' was initiated. Due to minor operating problems caused by the 10 day shutdown, several humidifier trips (i.e., loss of water feed) occurred.
- 8/29 The newly installed water spray lance dust blower system proved to be ineffective in keeping the lances free of hydrate buildups.
- 8/30 Inspection of the humidifier showed that water lance solids accumulations eventually interfered with the water sprays. This resulted in the deposition of a large amount of damp material on the humidifier floor directly in front of the atomizer array. The solids dropping off of the spray lances had built up high enough to intersect the water spray plume.
- 8/31 The solids distribution disks were reinstalled on the hydrated lime injection pipes. Two tests were conducted at high lime injection velocities (nominally 107 fps and 214 fps) to determine if the high discharge rates would break up lime particle agglomerates and better distribute the solids across the duct. The high velocity operations had no effect on the buildup of solids on the water spray lances. A third test using dried conveying air was also unsuccessful in controlling the deposition.
- 9/1 The solids injection pipes were extended through the water spray lance array. The solids discharged from the injection pipes in the plane of the water sprays. This modification was effective in eliminating water lance depositions.
- When testing began, the stack SO<sub>2</sub> analyzer failed so removal data was not obtained.
- 9/2 Boiler was taken off-line because of weekend power curtailment.  
to  
9/3

9/6 Conducted Coolside process tests at Ca/S = 1.0 and 2.0, Na/Ca = 0.0, and 20°F approach.

9/7 Coolside process operations were curtailed while the ash conveying system was repaired. The dump gate valve on the ash conveying system cyclone was sticking open.

Because of calibration drift problems with the existing stack gas analyzer, efforts were made to obtain a newer more advanced replacement unit. The new stack gas analyzer was operated in parallel with the original unit to allow comparison of results and gain confidence in its operation.

9/8 Ran tests with six and then with nine injector pipes in service. There was no difference in the SO<sub>2</sub> removals.

The atomizing air compressor shut down at 19:40 hrs due to a compressor thermocouple failure. Humidification was shut down until repairs could be made.

9/9 The system was off-line for the weekend.

to  
9/10

9/11 The Coolside process equipment was started up for a 48 hour test at Ca/S = 2.0, Na/Ca = 0.0, and 20°F approach.

9/12 At 17:45 hrs, the atomizing air flow control valve suddenly closed. The system operations were quickly recovered. The cause of the valve closing could not be determined.

9/13 The humidifier was shut down for inspection. The first third of the humidifier floor downstream of the atomizer array was clean. The second third of the floor area had some solids with a moist crust. The remaining floor area was covered with dry dust. Wall buildups were about 1" thick on the north wall and 1/2" to 1" thick on the

south wall downstream of the atomizer array. This material was soft and would slough readily.

Four atomizers in the center of the atomizer array had solids buildups on the nozzles and nacelles. Later inspections always showed that these atomizers were prone to deposition. The cause may have been due to insufficient shield air flow through the nacelles or to an unusual flue gas flow pattern around these atomizers. To reduce the localized solids deposition, the center hydrated lime injector was normally valved out of service.

During the afternoon and evening shifts, the Coolside process operations were reestablished but because of utility system dispatch projections, the equipment was shut down at 22:30 hrs.

9/14 The Coolside process equipment was brought back on-line at 08:48 hrs with Ca/S = 1.0.

Sodium additive (NaOH) feed was established for the first time. However, the continuous on-line sodium ion analyzer would not hold calibration. This analyzer was used to control the sodium additive supply to the system.

During the evening shift, the humidification system was shut down due to boiler going to minimum load.

9/15 Coolside process operations were reestablished at Ca/S = 2.0, Na/Ca = 0.185, and 20°F approach. When the sodium ion probe calibration was checked, the reading was 26 percent low. At 20:50 hrs, the system was shut down since the boiler was to be off-line for the weekend.

9/16  
to  
9/17 Off-line for weekend.

9/18 Before testing was resumed, the humidifier was inspected. Ninety percent of the atomizers were free of solids deposits. Only the four atomizers in the center of the atomizer array had severe deposits. Soft, friable wall scales were 0"-2" thick on the north wall and 0"-1" on the south wall. The roof scale deposits were spotty with many spalled areas. The floor debris was mostly dry dust with a small amount of crusty material from wall scales and nozzle deposits mixed in.

The first third of the floor area downstream of the atomizer array was clean (bare metal). The last two-thirds of the floor area had dust/solids buildups up to about 1" thick on top of the dust blower pipes. Between the blower pipes the dust laydowns tapered to the floor due to the cleaning action of the floor blowers.

The humidifier was closed up and testing commenced again in late afternoon.

The sodium additive feed system had to be taken off line because of incompatibility of the sample line filter element material with the caustic solution.

In preparation for switching to high-sulfur coal, the capacity of the hydrated lime feed system was tested by operating at high lime feed rates (i.e., Ca/S = 3.5 for the low-sulfur coal). ESP fields 1A and 1B shut down, apparently because of high hopper levels.

9/19 The test objective was to run for 24 hours with sodium additive feed. When it was determined that the on-line sodium ion analyzer would not hold calibration, the manufacturer of the instrument was contacted for recommendations. Because the ion probe was out of service, Coolside tests were conducted at an intermediate Ca/S ratio of 1.5.

9/20 The Coolside process equipment was shut down to allow plant operators time to clear high solids levels in ESP hoppers. A check of the material in the ESP hoppers showed that the material was dry.

The sodium ion probe used to measure additive concentration in the humidification water was recalibrated per the manufacturers recommendations. The unit still failed to hold calibration. To overcome this problem, the decision was made to install instrumentation to measure the flow rate of the concentrated additive feed. Near the end of the test program, the sodium ion probe was replaced with a conductivity probe which did function well.

- 9/21 Since a flow meter to measure the sodium additive feed could not be immediately installed, tests were conducted to establish the feed rate as a function of the feed control valve position. The concentrated additive feed rate was determined from the feed rate of humidification water and manually measured concentration of additive in the humidification water.
- 9/22 After completing the calibration of the sodium additive feed valve, a short Coolside process test was conducted with additive feed. The system was then shut down in preparation for the boiler being down during the weekend.
- 9/23 No testing on weekend.  
to  
9/24
- 9/25 Conducted test with boiler firing low-sulfur coal for following target  
to conditions:  $Ca/S = 1.0$ ,  $Na/Ca = 0.185$ , and  $20^{\circ}F$  approach.  
9/26
- 9/27 Conducted test for target conditions:  $Ca/S = 2.0$ ,  $Na/Ca = 0.185$ ,  $20^{\circ}F$  approach, and low-sulfur coal.
- 9/28 Conducted a short-term test at  $Ca/S$  ratio of 3.5 with sodium additive feed. The purpose was to check out both the hydrated lime feed and additive feed systems under conditions which simulated the use of high-sulfur coal in the boiler.

Lost electrical operation of ESP fields 1A, 1B, and 2B during the test.

At 17:31 hrs, the humidification water flow controller malfunctioned and too much water was fed to the system. The cause of the malfunction could not be determined.

9/29 To allow the ESP fields to recover energization, the Ca/S ratio was reduced to 0.5 and humidifier outlet temperature was increased from 142°F up to 200°F.

The humidifier was taken off-line for inspection and clean out prior to testing with high-sulfur coal.

The inspection showed that the four atomizers in the center of the array had the usual large solids buildups. The remaining atomizers were nearly clean or had very minor buildups. The first third of the humidifier floor downstream of the atomizers was clean (bare metal). For the remainder of the floor area, piles of solids built up to a depth of 2' to 3' on top of the dust blower pipes. Valleys formed between the blower pipes where the air blowing action was effective in clearing away the solids. Typical soft, dry, friable wall scales up to 3" thick formed on the side walls and roof of the humidifier. There was a large 4' deep pile of solids and scale material in the outlet turning vane section. This material tapered to the floor and did not extend in the outlet duct past the turning vanes. The water flow upset experienced on 9/28 was likely responsible for much of the turning vane solids accumulations.

ESP operating problems were discussed with Ohio Edison personnel. The consensus was to increase rapping intensity and frequency.

9/30  
to  
10/1

System was down over weekend for humidifier cleanup.

10/2 Began operating the Coolside process for the first time with high-sulfur coal feed to the boiler. To minimize the potential for ESP operating problems, started testing at a low hydrated lime feed rate. Target conditions were: Ca/S = 1.0, Na/Ca = 0.185, and 20°F approach.

10/3 The hydrated lime feed rate was increased to obtain a Ca/S ratio of 2.0. The power level to the front ESP fields decreased somewhat. ESP volt-current (V-I) data were obtained to characterize the operation.

10/4 The hydrated lime feed rate was decreased to achieve a Ca/S ratio of to 1.6.

10/6

The sodium additive feed pump developed a seal leak which allowed flush water to leak into the process. This diluted the stored concentrated feed somewhat since the pump discharge recirculates back to the feed storage tank.

10/7 The humidifier was shut down for the weekend and for inspection. Found typical atomizer deposits, wall scales, and a clean floor for the first third of the humidifier length. Powdery, dry dust piles had again built to a height of 2'-3' above each of the floor blowers downstream of the bare floor area. A large accumulation of dust and scale debris had formed in the outlet turning vanes. This restricted the outlet gas flow area by about 50 percent. The decision was made to clean out only the turning vane area to just past the outlet louvered damper. The short duct area from the turning vanes to the outlet louvered damper was cleaned because the 1"-2" deep floor rubble had kept the damper from being completely closed during the shutdown.

10/10 The Coolside process equipment was put into service. To reduce the solids loading to the ESP and because of a high back pressure on the 'A' hydrate feeder, the hydrated lime feed rate was reduced to provide a Ca/S ratio of 1.4. On the following day, the 'A' feeder problem was found to be caused by a lime deposit at the pneumatic conveying system solids pickup point.

10/11 Continued operation at target values of Ca/S = 1.4, Na/Ca = 0.185, and 20°F approach. To keep power on the front ESP fields (1A, 1B, 2A, and 2B), the transformer/rectifier (TR) set controllers were adjusted to provide intermittent energization with a 10 percent on - 90 percent off cycle.

Attempts were made by Consol R&D to measure in situ the ESP inlet particulate resistivity by the point plane method. Solids were collected rapidly in the resistivity device. However, a good resistivity measurement could not be obtained. The performance of the instrument indicated that particle resistivities were not high.

10/12 The Ca/S ratio was increased from 1.4 to 2.0.

10/13 The boiler was taken off line for several hours because of feed water pump problems. When the boiler was back in service, Coolside test conditions were reestablished at: Ca/S = 2.0, Na/Ca = 0.185, and 20°F approach. During the evening shift, the equipment was shut down in preparation for the boiler being off-line during the weekend.

Inspection of the humidifier showed the turning vanes to be relatively free of deposits. However, there was a large 4' deep pile of solids at about one-third of the humidifier length downstream of the water spray atomizer array. In other areas, the floor dust piles were normal in appearance. The dust piles were leveled out to produce a 2' thick dust layer across the floor and the atomizer array was cleaned by wire brushing. Large deposits were again removed from the four atomizers in the center of the 100 atomizer array.

10/14  
to The system was down because of low boiler load over the weekend.

10/15

10/16 The humidifier was brought on-line at target condition of Ca/S = 2.0, Na/Ca = 0.185, and 20°F approach.

10/17 The ash conveying line from the ESP hoppers to the ash silo plugged due to rain water leakage at a pipe coupled joint. This line was replaced with an all welded line later in the test program. When the line was being cleaned, several ESP hoppers developed high solids levels. The Ca/S ratio was reduced from 2.0 to 1.0 to reduce the ESP inlet particulate loading.

At 14:00 hrs, the humidifier water flow control valve again opened for no know reason. This problem occurred occasionally throughout the test program. The cause could never be found. Voltage spikes were at one time thought to be responsible. Monitoring of the voltage supply failed to substantiate this as the cause.

10/18 The Ca/S ratio was increased to 2.0.

10/19 Because the humidifier outlet thermocouples (TCs) did not appear to be operating properly (i.e., the humidifier outlet TCs indicated higher temperatures than the ESP inlet TCs), the humidifier temperature control was switched to the ESP inlet TC array. This was an adequate means of controlling the humidification water rate since humidifier flue gas bypass was not being used during these tests.

At 09:57 hrs, the Ca/S ratio was reduced from 2.0 to 1.0 to allow operations time to clear the high ESP ash hopper levels.

The system was shut down later in the day when it was determined that there was a large solids buildup at the humidifier outlet.

10/20 Inspection of the humidifier showed that the atomizer array was in good condition except the usual heavy deposits on the four center atomizers. Dust and debris levels were up to 4' deep on the humidifier floor. A massive 8' high solids deposit at the outlet reduced the gas flow area by 75 percent. Much of this deposit was made of hard material which had been wet. Three of the five outlet TCs were buried. This explained the lack of response of these TCs noted during the run.

A review of the process operations data indicated that the large amount of deposition resulted from operating with too much gas flow through the humidifier. For extended periods, all the gas flow was through the humidifier, the boiler load was high at 90-100 MWs and because of air heater leaks the excess air in-leakage was also high. This reduced the available drying time. Additionally, because of the high water rates required for humidification, the atomization air/liquid ratio was below 0.4 lb/lb which is less than the design ratio of 0.45 lb/lb. This decreased the atomizer performance and increased the spray droplet size, further increasing the drying time requirements.

Operating procedures were revised to address the aforementioned problems. The major changes included: limiting the flue gas flow through the humidifier, maintaining air/liquid ratios of 0.45 lb/lb or greater, and maintaining all humidifier outlet TCs including unshielded TCs at temperature approaches to saturation of 10°F or greater.

10/21 The boiler feed was switched back to low-sulfur coal to insure that the utility SO<sub>2</sub> emissions remained in compliance while the Coalside process was down for humidifier cleanout.

10/22 Continued with the humidifier cleanout.

10/23 Humidifier cleanout was completed.

The humidifier outlet TC supports and shields were modified to reduce the area on which deposits could form.

10/24 The boiler was down most of the day because of problems with the house service water pumps.

10/25 Began bringing the humidifier back on-line. Testing indicated that the humidifier would provide adequate drying time for 900,000 lb/hr of flue gas flow.

- 10/26 Established stable operating conditions at: Ca/S = 2.0, Na/Ca = 0.185, and 20°F approach. At 16:30 hrs, the humidification feed water valve closed for no apparent reason. The system was quickly recovered from the upset.
- 10/27 The humidifier was shut down due to excessive conveying air leakage past the rotary valves which are below the hydrated lime gravimetric feeders.
- Inspection of the humidifier showed it to be in good condition.
- 10/28  
to  
10/29 Repairs were made to the two hydrated lime feed system rotary valves.
- 10/30 Began bringing the Coolside process equipment back into service.  
to  
10/31 Tests were made to check out the effect of the feeder repairs on the operation of the hydrated lime feeding system.
- 11/1 Established operating conditions at: Ca/S = 2.0, Na/Ca = 0.24, and 20°F approach with the boiler firing low-sulfur coal.
- 11/2 A high feed rate test was conducted on the 'B' hydrated lime feeder. Repair of the boiler system water pumps required that the Coolside equipment be shut down.
- 11/3 Inspection of the humidifier showed the system to be in good condition with very little material on the floor. The thickest dust layers were only 1-1/2" to 2" above the dust blower pipes. Only one atomizer in the center of the array had a large solids deposit. The outlet turning vanes had deposits of lime dust and crusty material from 1" to 7" thick on the sides facing the gas flow. This indicates that some moist solids still impact the vanes in the turn. As always, there were no deposits but only a soft dust layer adhering to the back side of the turning vanes. The humidifier was cleaned out during the boiler shutdown.

- 11/4 No testing because of boiler repairs and problems with the boiler  
to house water intake system.  
11/8
- 11/9 While the boiler was down, two Teflon® sheets were attached to the humidifier outlet turning vanes to determine if non-stick surface materials would be effective in controlling solids depositions.
- 11/10 Because of plugging problems experienced during the last run, the three bottom hydrated lime injectors were valved out of service.
- The plant began placing the boiler back in service during the afternoon shift. Because of the long down time, the start-up was protracted.
- 11/11 Began start-up of the Coolside process equipment with the boiler firing high-sulfur coal. To improve water atomization, the minimum air/liquid ratio was increased to 0.55 lb/lb.
- 11/12 Two of the hydrated lime injector pipes plugged. One injector was cleared. Also had a problem with the pocket vent system of the rotary feeder located below the 'B' hydrated lime gravimetric feeder. Plugging of the vent line caused a back pressure on the gravimetric feeder which limited the lime feed to less than 6,000 lb/hr.
- 11/13 At 09:15 hrs, shut down the 'B' hydrated lime feeder for a half hour to clear the rotary air seal feeder vent line.
- Shut down humidification and hydrate feeding at 14:30 hrs because of high solids levels in four ESP hoppers. The inability to empty the ESP hoppers was caused by plastic debris lodging in the hydro eductor used to pull the ash system vacuum.
- The boiler was taken off line to repair a feed water pump.

- 11/14 The Coolside process was started up at: Ca/S = 2.0, Na/Ca = 0.24, and 20°F approach.
- 11/15 The humidifier was shut down because of high load cell readings. A large amount of settled solids was inside of the humidifier. The solids accumulation was likely the result of poor atomizer performance due to foreign debris being trapped in or passing through the nozzles. Five of the atomization nozzles were completely plugged with a material which appeared to be fish pulp. This material probably came from the plant's water intake mechanical strainers.
- 11/16 The humidifier remained off line while the atomization feed water  
to system was repiped to eliminate leaking screw fittings. This was  
11/20 necessary because the water was caustic when sodium additive was fed.
- 11/21 The humidification system was operated to evaporate caustic water in the water storage tank which was left over from the previous test. The humidifier was then shut down to inspect the atomizers. A qualitative test know as a "wrist" test was preformed to check the atomizer performance. The test consists of feeling with one's hand the atomizer spray. If no droplets can be felt in the spray, the atomizer is judged to be operating properly. The results of the test indicated that the performance of about 13 percent of the atomizers had significantly deteriorated.
- 11/23 Off line because of Thanksgiving holiday.  
to  
11/25
- 11/26 The plugged hydrated lime injectors were cleared. The system was then set up to operate with five of the nine injectors in service.
- 11/27 Began Coolside process equipment start-up.

- 11/28 Began the start-up of the ash recycle system but had to shut down at 18:30 hrs. The hydrated lime feeders continually tripped off line due to lime material packing tightly around the gravimetric feed screws.
- 11/29 The hydrated lime feeders were cleaned out and special feed rate tests were conducted to determine system operability and the maximum deliverable feed rates. Achieved 7,800 lb/hr and 9,800 lb/hr feed rates from feeders 'A' and 'B', respectively.
- 11/30 Began feeding recycle solids to the unit for the first time to increase the sorbent utilization. Operating conditions were: Ca/S = 1.0, Na/Ca = 0.185, 20°F approach for all humidifier outlet TCs, and 4,000-5,000 lb/hr of recycle solids.
- The sodium hydroxide additive feed pump developed a seal water leak which caused the storage tank to overflow into the catch basin around the tank. The seal water was shut off to prevent further overflow.
- 12/1 The sodium hydroxide additive feed pump was shutdown at 10:31 hrs for repair. Coolside process operations continued without additive makeup.
- At 20:30 hrs, the recycle solids feed system was shutdown because of high levels in four of the 12 ESP hoppers. The system was restarted at 01:00 hrs on 12/2.
- At 21:00 hours, a leak was discovered in the stack oxygen analyzer system.
- 12/2 Operations continued although all stack gas analyzers were out of service.
- 12/3 Held operating conditions of Ca/S = 1.0, Na/Ca = 0.0, 20°F approach, and 4,000-5,000 lb/hr of recycle solids while gas analyzers were under repair.

12/4 By 10:30 hrs, the stack SO<sub>2</sub> had been repaired. Continued the test with recycle solids feed.

12/5 Because of high ESP hopper levels, the recycle solids feed was discontinued at 14:45 hrs.

12/6 Continued operating the Coolside process equipment at Ca/S = 1.5, Na/Ca = 0.185, 20°F approach, and no recycle solids feed.

Shut down the Coolside process equipment at 18:15 hrs to allow plant operations and a manufacturer's representative to inspect the ash conveying system. Inspection of the ash system baghouse did not reveal evidence of moisture caused problems.

12/7 While the ash conveying system was out of service, the humidifier was inspected. The unit was in good condition. The floor material between the atomizer array and the outlet was mostly dust and wall scale debris. There was relatively little of this material. Along the humidifier centerline the material depth ranged from 1/2" to 9-1/2".

Along the walls, the material was 12"-18" deep due to the spalling of wall scales. The south wall was mostly bare metal because the scale had spalled. The north wall had 1"-2" thick loose scale.

The outlet turning vanes had crusty deposits on the front sides which were exposed to gas flow impaction. These deposits were typically 1-1/2" to 2" thick. In one area the deposit was about 5" thick. The Teflon<sup>®</sup> sheet, which had been attached to a turning vane pipe support, was free of scale. The Teflon<sup>®</sup> covered area of the center turning vane was partially covered with a deposit which readily sloughed off when touched. However, thirteen of the water atomizers had large deposits on the nozzles or nacelles which likely affected the atomizer performance.

12/8 Cleaned out the humidifier for the next test.

12/9 During most of the day, problems were experienced with the atomizing air compressor operation. The problems were caused by loose attachment of the vent valve instrumentation.

At 17:20 hrs, operations began bring the humidifier on-line and establishing the following conditions: Ca/S = 1.5, Na/Ca = 0.185, no solids recycle, and 20°F approach consistent with no humidifier outlet TC being below this approach.

12/10 For the following week, the plant agreed to limit the boiler load to 70 MW maximum generation during Coolside process test periods. This allowed processing all of the flue gases through the humidifier and low temperature operation of the ESP to increase SO<sub>2</sub> removals.

12/11 Recycle solids injection was reestablished at 2,000-3,000 lb/hr. One recycle solids injector plugged.

The sodium additive feed rate was reduced by one half to conserve reagent pending the arrival of a tanker load to resupply the system.

12/12 Because of ash conveying system problems with the hydro eductor unit and a plugged ash line, the plant requested that the Coolside process be shut down to reduce the particulate load to the ESP. The system was down by 06:07 hrs.

12/13 There was no testing because of ash system repairs.

12/14 The Coolside process equipment was started up at 00:10 hrs, but the sodium additive could not be fed because the feed pump discharge pressure was low.

At 08:15 hrs, momentarily lost the atomizing air flow due to a controller malfunction which shut the pressure control valve. The system quickly recovered from the upset.

Shut down the Coolside process equipment at 20:00 hrs in preparation for a scheduled boiler shutdown at 23:00 hrs.

12/15 Inspection of the ESP revealed the presence of hotdog-shaped wire deposits mainly in fields 1A, 1B, 2A, 2B, 3A, and 3B. The front fields had the largest amount of deposits. The deposits formed on wires which were furthest frame rapping points. Wires which were close to rapping points were free of deposits. This indicates that installation of more rappers or more intense rapping would be effective in controlling wire deposit formation. The ESP plates appeared to be in good condition.

The floor debris in the humidifier was 2'-2.5' deep. A close examination of the atomizer discharge ports revealed that many were not round as originally installed but were egg-shaped. This indicates that significant wear had occurred.

12/19 The Coolside process had a false start-up when the boiler was brought back on-line and then taken off-line to fix a leaking blowdown valve weld.

12/20 The boiler remained down because of intake water system problems.

12/21 Could not establish Coolside process operations because the humidification water supply line had frozen.

12/22 System was down due to holidays and vacations.

to

1/1

1990

1/2 The motor for the hydrated lime conveying air compressor was found to have shorted out. It was removed and sent out for rewinding.

1/5 The hydrated lime conveying air compressor motor was reinstalled and Coolside process equipment was started up.

1/6 At 01:45 hrs, the Coolside process equipment was shut down due to problems with the ash conveying system. The air lock system which allows the ash conveying system cyclone to dump solids into the ash silo plugged. The problem was traced to a plugged aeration air distribution ring.

The hydrated lime conveying air compressor could not be operated because of an electronic module failure.

1/7 The system remained down due to problems with the hydrated lime conveying air compressor control.

1/8 The electronics problem with the hydrated lime feed system was manually bypassed and the Coolside process equipment started up.

At 17:32 hrs, another humidification water flow control upset occurred. The water flow first dropped off and then the flow control valve opened fully for no apparent reason. The problem appeared to be an electronics glitch.

1/9 The hydrated lime conveying air compressor control problem was fixed. The Coolside process was operated to empty out the inventory of hydrated lime 'A' in preparation for testing with hydrated lime 'G'.

1/10 The stack gas analyzers failed due to a corroded stack probe. After the probe was repaired, the analyzers still failed to operate properly.

A repaired sodium additive (sodium hydroxide) pump was reinstalled. However, sodium hydroxide was not fed into the humidification water because meaningful data could not be obtained without the stack gas analyzers operating.

At 17:40 hrs, an automatic valve in a no longer used differential pressure control system for the atomizing air supply closed due to an instrumentation problem. This shut off the high pressure atomizing air supply to one-sixth of the water sprays. Without this air, unatomized water was fed directly into the humidifier for about 18 minutes. Based on the response of floor thermocouples, water covered the floor up to 44' downstream of the atomizers. The water supply and lime feed to the humidifier were shut off to allow hot flue gas evaporation of the water already in the humidifier.

Hydrated lime and water flow to the humidifier were reestablished at 21:00 hrs. The Ca/S ratio was increased to 3.0 to more rapidly deplete the inventory of hydrated lime 'A'.

1/11 The stack gas analyzers remained out of service. The Coolside process equipment was operated just to deplete the hydrate 'A' inventory.

1/12 Repair of the stack gas analyzers continued.

At 09:33 hrs, the humidification water flow control valve opened fully for no identifiable reason. This dumped a large amount of excess water into the humidifier for 27 minutes. The water flow was shut down automatically when the outlet temperature reached 129°F. Because 30 percent of the hot flue gases were being bypassed around the humidifier the ESP was not exposed to a fully saturated flue gas. The humidifier outlet temperature was raised to 220°F to help dry out the system.

At 20:00 hrs, began shutting down the Coolside process to allow inspection of the humidifier.

1/13 Considering the process upsets experienced during the week, the humidifier was in relatively good condition. The solids on the humidifier floor were 1'-1.5' deep. This material was dense and clumped together when squeezed in the hand, indicating a high moisture content. The outlet turning vane deposits were 3"-4" thick. Only one

of the four atomizers in the center of the atomizer array which normally have tulip-shaped deposits had a large deposit. The other three atomizers had only small buildup.

The Coolside process equipment remained off line until the gas analyzers were repaired.

1/16 Start-up of the Coolside process equipment began at 09:30 hrs. The following test conditions were established: Ca/S = 1.0, Na/Ca = 0.0, 20°F approach, and boiler firing low-sulfur coal.

1/17 The Ca/S ratio was raised to 2.0 at 08:45 hrs.

1/18 Reduced the Ca/S ratio to 1.0 at 09:05 hrs.

Attempted to operate the sodium additive feed system, but the caustic feed line was frozen. Attempts to thaw the system were unsuccessful.

1/19 The sodium additive feed system was thawed out at 23:15 hrs by using an arc welder power supply to heat the piping. The Na/Ca ratio was established at 0.185.

1/20 At 11:00 hrs, a humidification water isolation valve closed unexpectedly. The system operation was quickly recovered following the upset.

1/21 The hydrated lime rate to the system was kept low so that the plant maintenance could replace the bags in the ash system baghouse.

At 14:00 hrs, the plant operations requested shut down of the Coolside process to allow repair of the hydro eductor water sluice line to the ash pond. The plant wanted to minimize the accumulation of ash in the system while repairs were made.

Hydrated lime feed to the system was reestablished at 23:33 hrs.

1/22 The following conditions were established: Ca/S = 2.0, Na/Ca = 0.185, and 20°F approach.

1/23 Due to a low pump discharge pressure, the sodium addition was lost. Feed was reestablished after the pump impeller clearance was adjusted at 17:15 hrs.

1/24 Again lost the sodium additive feed to the system at 06:50 hrs. The pump impeller clearance had to be readjusted.

At 14:05 hrs, the Ca/S ratio was set to 1.0.

1/25 The boiler was switched to firing high-sulfur coal.

Due to a plugged strainer, the humidification water flow was momentarily lost. This was corrected by switching to a parallel strainer.

The sodium additive pump seal was repacked because of excessive sodium hydroxide leaks.

The Ca/S ratio was increased to 1.5 and the Na/Ca ratio was established at 0.185.

1/26 Due to a pump motor trip, the sodium additive feed was lost at 05:44 hrs.

At 18:00 hrs, high opacities were experienced when the boiler load changed from 75 MW to 103 MW. Opacity spikes to 7.5 percent were noted (limit <10 percent). To avoid opacity excursions, the Ca/S ratio was reduced to 1.0.

1/27 The Ca/S ratio was increased to 1.5.

Sodium additive feed was discontinued at 08:15 hrs.

The Ca/S ratio was reduced to 0.5 at 15:00 hrs to maintain opacity. Process operations had resulted in ESP controllers dropping the power to zero on fields 1A, 1B, 2A, and 3A. Field 2B was out of service for reasons not associated with the testing and the power to field 3B was dropping.

Because the boiler was to be taken off-line, the Coolside process equipment was shut down at 20:45 hrs.

1/28 Inspection of the humidifier showed that the floor was clean for about half the length downstream of the atomizer array. From this point on to the outlet, the maximum depth of solids on top of the floor blower pipes was about 3'. Several atomizers had soft ball-sized, tulip-shaped deposits. The humidifier was not cleaned out during this shutdown.

The Coolside process equipment was restarted at 18:45 hrs.

Rapping of the ESP fields while the boiler was down had restored performance.

1/29 Because the hydrated lime feeders would not automatically refill, the humidification water feed was stopped from 03:40-08:40 hrs.

Additional packing was added to the sodium hydroxide feed pump to stop shaft seal leaks.

Coolside process operations ceased at 16:10 hrs because the boiler was taken off line.

1/30 During the boiler outage, the humidifier was again inspected. This inspection was after approximately two weeks of Coolside process operations without humidifier clean out. The floor debris was dry. The solids accumulation on top of the floor blower pipe which was midway in the humidifier was the largest and was approximately 3' high. The deposit consisted of dust with a top crust. Apparently,

the material had built up high enough to intersect the water spray plume traveling down the humidifier during operations. This would explain the top crust material. The floor blowers had been effective in preventing solids accumulation between the blower pipes.

Wall scales were nominally only about 1/4" thick.

The outlet turning vanes had 3"-4" thick crusty scale deposits on the sides facing the gas flow and very soft 1"-2" thick dust scales on the back sides. Both types of scale were easily knocked off the vanes. The sloughing of these deposits during the operation had produced a 2' deep floor rubble pile in the turning vane area.

A thin, crusty scale tenuously adhered to the test Teflon<sup>®</sup> sheet on the center turning vane. When lightly touched, this material fell off the sheet. The testing indicates that non-stick surfacing of the turning vanes can help limit solids deposition due to the impact of still moist particles in this area of the humidifier.

The atomizer array was in good condition. Only four atomizers scattered around the 100 atomizer array had large tulips.

- 2/1 The boiler was started up again firing high-sulfur coal. Once stable boiler operation was achieved, the Coolside process equipment was brought back on-line. Target conditions were: Ca/S = 1.5, Na/Ca = 0.0, and 20°F approach.
- 2/2 Sodium additive feed was established at 15:45 hrs after the caustic pump and feed line were thawed. The target feed level was Na/Ca = 0.185.
- 2/3 Operations continued. The coal sulfur level began decreasing due to the change over to low-sulfur coal.

The Ca/S ratio was reduced from 1.5 to 1.0 at 15:45 hrs.

The Coolside process was shut down at 20:20 hrs due to the boiler being taken off-line.

2/4 The boiler was off-line all day.

The sodium additive feed pump was repacked to eliminate shaft seal leaks.

2/5 The boiler was brought back on-line at 07:00 hrs.

Began Coolside tests using hydrated lime G and solids recycle to increase the sorbent utilization. Target conditions were  $Ca/S = 0.75$ ,  $Na/Ca = 0.185$ , 20°F approach for closest approach of any humidifier outlet thermocouple, and recycle solids rate of 4,000-8,000 lb/hr.

2/6 Maintained Coolside process operating conditions. DOE/PETC sponsored ESP rapping loss tests were initiated. During these tests, ESP fields 1A and 1B were purposefully turned off to decrease the effective size of the ESP.

2/7 Maintained Coolside process operating conditions.

2/8 Two recycle solids injector lines developed plugs which were cleared.

2/9 Recycle solids flow was lost through five of the nine feed lines from the distribution bottle. Clinker material in the recycled ash solids plugged the 2" diameter feed line block valves. The lines were cleared by opening and closing the block valves. This broke up and crushed the clinker material which was lodged in the valves. A coarse trash screen before the pneumatic pickup point at the ash silo would eliminate this problem in a commercial system.

2/9 Had problems maintaining ESP energization during the day. Fields 1A and 1B had been shut down for special rapping loss tests. When turned on, these fields would not operate since voltages remained at zero.

Only fields 2A, 3A, and 6AB operated well throughout the day. Despite this, the opacity generally remained below 3 percent.

2/10 At 12:40 hrs, the sodium additive feed was lost due to low pump discharge pressure. Flow was recovered by nearly closing off the pump recirculation line to the feed tank. At 16:49 hrs, the sodium additive pump tripped off. The water flushed shaft seal also leaked excessively.

2/11 Conducted a short 5.5 hour test to determine the SO<sub>2</sub> removal due to recycle solids feed only.

Attempts to add packing to the sodium additive pump failed to stop shaft leakage. The pump required a complete repacking and lantern ring replacement.

Began to find leaks in the rubber hoses connecting the recycle solids distribution bottle to the humidifier duct solids injector pipes. The abrasive clinker and ash material transferred through these lines had caused erosion. Taping the hoses temporarily fixed the problem.

The recycle solids feed was lost for a while at 16:30 hrs due to clinker material jamming the feed valve at the pneumatic conveying line inlet.

2/12 The recycle feed system was down from 00:55 hrs to 01:21 hrs because of clinker material jamming the rotary feed valve.

The recycle solids feed lines continually plugged at the distribution bottle. The recycle solids system was shut down at 07:50 hrs when a leak developed in a pipe nipple on the distribution bottle.

All plugged recycle solids valves on the distribution bottle were cleared and the eroded hoses and the 2" diameter eroded pipe nipple were replaced.

The sodium additive feed pump was completely repacked and the impeller repositioned. This corrected the low discharge pressure problem. The pump was placed back in service at 20:36 hrs.

2/13 Coolside operating conditions were maintained at: Ca/S = 0.75, Na/Ca = 0.185, 20°F approach, and 4,000-8,000 lb/hr recycle solids.

2/14 Maintained operating conditions.

2/15 The hydrated lime feed was lost for about 1.5 hours beginning at 02:00 hrs. The problem was apparently due to the plugging of a feeder vent line.

Between 11:00 hrs and 11:37, hrs the hydrated lime feed was again off while the feeder vent system was unplugged.

Between 11:05 hrs and 11:55 hrs, the recycle solids feed was lost due to a shutdown of the transport air compressor.

At 14:49 hrs, the sodium additive feed pump impeller clearance was readjusted because of low discharge pressure. The pump was back in service at 16:38 hrs.

2/16 Conducted a five hour test at a closer temperature approach. Test conditions were: Ca/S = 0.75, Na/Ca = 0.185, 10°F approach for the closest approach of any humidifier outlet thermocouple reading, and 8,000 lb/hr of recycle solids feed. The actual temperature approach was 17°F based on averaging the five humidifier outlet thermocouples.

Conducted a second four-hour test at the closer temperature approach but with only recycle solids feed (no hydrated lime or sodium additive fed). The average temperature approach was 19°F.

Shut down the Coolside process equipment at 09:57 hrs. Obtained comparison of the humidifier inlet and stack gas SO<sub>2</sub> analyzer measure-

ments without system SO<sub>2</sub> removal. When corrected to zero moisture and excess air, the analyzer measurements were in agreement.

Inspected the humidifier. The humidifier was found to be in good condition after eleven days of continuous operation at a 20°F approach. The soft wall scales were mostly 1/4" to 1/2" thick.

There was little material on the humidifier floor. For the full length of the humidifier, the areas between the dust blower pipes had only a light dust coating or were bare metal. The dust material on top of the blower pipes was only a few inches thick but had a hard crust. The outlet turning vanes had a 2" thick crust on the front sides which were exposed to the gas flow. The turning vane scale deposits were harder than material normally found after a run. This may have been due to the close temperature approach operating conditions established at the end of the run. The solids in these areas may have become wetted during the close approach test and then dried out and hardened when the humidification water was shut off for several hours before the equipment was finally shut down.

The louvered damper which is just downstream of the humidifier outlet turning vanes was coated with a thin hard scale. The damper could not be closed at the end of the run because the scale had bound the closing lever mechanism. No problems were experienced with the operation of the humidifier inlet and outlet guillotine dampers during any of the tests.

The atomizer array was in good condition. Only seven atomizers had large tulip deposits. As usual, three of the deposit-coated atomizers were in the center of the array. Simple mechanical devices such as retractable sootblowers, mechanical scrapers, or brushes could be used commercially to control these deposits.

## APPENDIX D. COOLSIDE AND WET LIMESTONE PROCESS MODEL PRINTOUTS

The attached computer model printouts are designed to show the model equipment sizing options and should not be considered to represent optimized process design configurations. For the Coolside process, the printout shows purchases in nearly all Coolside model equipment categories. This was designed to demonstrate the Coolside model capabilities.

## APPENDIX D. COMPUTER MODEL OUTPUTS

### TABLE D-1

#### ECONOMICS SUMMARY -- UNOPTIMIZED COOLSIDE PROCESS WITH ON-SITE QUICKLINE HYDRATION

28-Jun-91C O N S O L R & D      GENERIC--250 MW -- Pgh & Coal -- 1.5% S -- Coolside

#### ASSUMPTIONS

Unit	GENERIC--250 MW		No. Units		1
Fuel	Pgh & Coal -- 1.5% S				
MW, Gross	262		MW, Net	249.6	246.9
Net Heat Rate	9,507	9,609	Aux power	4.89%	5.90%
% Excess Air	40.0		Turb H R	7,861	[C]
Air Temp	60		Spare Hyd	0%	CO
F G Exit Temp	304		Reheat	0 deg	Manuf
SO2 Removal, %	70.00		Cap Fact	65.00%	Rad
Fresh Ca/S Ratio	1.33		Base Year	1990	
Int. Ca/S Ratio	2.38				
Lb H2O/lb bdair	0.0069		Humid Width	27.07 ft	
flyash/ash	80.00%		Unloading	1 0=Truck 1=Rail	
Plant Loc fac	1.06		Retrofit	1 0=New 1=Retrofit	
			ID Fan Suct	-14.5 in wg	
			Ca Util, cal	24.23	
			Ca Util, int	24.23	

Compositions, weight % dry basis			wt assuming		
Coal			Lime, wt%	Hydrate	100 lb BL
H	4.93		CaO	91.00	Ca(OH)2 120.23
C	79.89		MgO	0.00	Mg(OH)2 0.00
H + Cl	1.52		Inerts	9.00	Inerts 9.00
O	4.03				
S	1.59				
Ash	8.04				
HHV	14,180	14,427 Btu/lb			
Stoic Air	10.71				
Moist	5.50				

#### RESULTS

Efficiency	86.76%	SO2 Removal	70.00%
Net Heat Rate	9,527	SO2 Emissions	0.67 lb/MWh
O2, dry	6.28%		4,532 ton/yr
Flue Gas, MSCFH	574.6		0.80 tph
,MACFH	879.3	-8.5 M <sup>3</sup> H2O	
Particulate loading	10.46 lb/MMBtu	within system (includes recycle)	
	3.293 g/scf	25.00 lb/MMBtu	
		7.87 gr/scf	12.04 gr/scf
Flue Gas Composition, mol % or ppm			
	Dry	Moist	Particulate, weight %
CO	41	39	[C]
CO2	12.90%	12.11%	Ash+Inert
H2O	0.00%	6.16%	89.64%
N2	80.72%	75.75%	
O2	6.28%	5.89%	
SO2	972	912	
Mol Wt.	30.37	29.61	

(continued)



# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-1 (continued)

28-Jun-91C O N S O L R & D      GENERIC--250 MW -- Pgh B Coal -- 1.5% S -- Coolside

<p>Plant                    GENERIC--250 MW  Capacity                249.58 MW net ex FGD                    246.92 MW net incl. FGD  Coal                    Pgh B Coal -- 1.5% S  XS as received        1.50 %  HHV as rec.            13,400 Btu/lb  feed rate,base        88.53 ton/hr  Capacity Factor        65%      Heat Rate:    9,507 ex FGD    9,609 incl FGD  Retrofit Diff.        MEDIUM Site Access                    MEDIUM Undergrd Obst</p>	<p>Reagent Preparation  -----  90 Max Silo Diam., ft  108 Max Silo Ht., ft</p> <p>Fan  -----  2 Number/unit  2.0 Pres Drop, in. H2O</p> <p>-----  Power Consumption  -----  Reagent Prep &amp; Handling            356 kW  Flue Gas Handling                    22 kW  Humidifier                            2,349 kW  Waste Disposal and Recycle        97 kW  Particulate Removal                0 kW  Waste Handling                        104 kW  ID Fan Differential                  (274)kW  -----  Total                                    2,654 kW  -----  total</p>	<p>-----  Humidification  -----  20.0 Humid Inlet Vel, fps  0.19 Na/Ca Molar Ratio  1.33 Ca/S Mol Ratio, Fresh Sorbe  70.0 Design Removal  20 Humid Approach Temp, deg F  0.50 Lb Air/Lb H2O                    30  75.0 mol% CaSO3 in waste  270 No. Atomizers  3 Humid Res Time  1 No. Humidifiers  0.025 Pressur Drop, " H2O</p> <p>Duct Runs  -----  50.0 Duct Velocity, fps</p> <p>-----  Waste Handling System  -----  No Upgrade  99.00 % Solids in Waste  Air Compressor  -----  150 Discharge PSIG  1 No. Operating  0 No. Spares  4 No. of Stages  90 Interstage Air Temp  3 Interstage Delta P, PSI</p> <p>-----  Particulate Removal  -----  0.100 Particulate Limit, lb/MMBtu  99.60 ESP Efficiency, %  Chimney  -----  No Costs</p>
<p>fresh Ca/S mol rat    1.33  Wet bulb temp        120.9 F  Humid Temp            140.9 F  Water added           0.0401 lb/lb fg  216.3 gpm  11.85 MSCFM  Scrubbed Flue Gas    634 MSCFM  763 MACFM  Bypass Flow           0.0X  0 MACFM  Mix T<sub>a</sub>                140.9 deg F  FG Cp, estimated     0.26  Reheat Req            0.00 deg F  0.0 MMBtu/hr Steam  No. of Humidifiers    1 operating  Sorbent                3.40 tph Lime</p>	<p>-----  Reagent Preparation  -----  90 Max Silo Diam., ft  108 Max Silo Ht., ft</p> <p>Fan  -----  2 Number/unit  2.0 Pres Drop, in. H2O</p> <p>-----  Power Consumption  -----  Reagent Prep &amp; Handling            356 kW  Flue Gas Handling                    22 kW  Humidifier                            2,349 kW  Waste Disposal and Recycle        97 kW  Particulate Removal                0 kW  Waste Handling                        104 kW  ID Fan Differential                  (274)kW  -----  Total                                    2,654 kW  -----  total</p>	<p>-----  Humidification  -----  20.0 Humid Inlet Vel, fps  0.19 Na/Ca Molar Ratio  1.33 Ca/S Mol Ratio, Fresh Sorbe  70.0 Design Removal  20 Humid Approach Temp, deg F  0.50 Lb Air/Lb H2O                    30  75.0 mol% CaSO3 in waste  270 No. Atomizers  3 Humid Res Time  1 No. Humidifiers  0.025 Pressur Drop, " H2O</p> <p>Duct Runs  -----  50.0 Duct Velocity, fps</p> <p>-----  Waste Handling System  -----  No Upgrade  99.00 % Solids in Waste  Air Compressor  -----  150 Discharge PSIG  1 No. Operating  0 No. Spares  4 No. of Stages  90 Interstage Air Temp  3 Interstage Delta P, PSI</p> <p>-----  Particulate Removal  -----  0.100 Particulate Limit, lb/MMBtu  99.60 ESP Efficiency, %  Chimney  -----  No Costs</p>
<p>Soda Ash               0.54 tph  SO2 Removed           1.86 tph  10,575 ton/yr  70X  SO2 emission          0.80 tph  4,532 ton/yr  0.671 lb/MMBtu  Solid Waste            6.23 tph dry ash free                    Wet Solid Waste    12.49 tph  12.36 tph dry waste  Power                   2,654 kW</p>	<p>(Ca/S)<sub>f</sub>                1.33  (Ca/S)<sub>int</sub>              2.38  (Na/Ca)<sub>f</sub>               0.19  (Na/Ca)<sub>int</sub>            0.25</p>	<p>-----  Particulate Removal  -----  0.100 Particulate Limit, lb/MMBtu  99.60 ESP Efficiency, %  Chimney  -----  No Costs</p>
<p>Max Allowable Side    29.66 tph  Rec. Ratio, R/W        1.39  Recycle Solids        17.18 tph  Actual Emissions      0.10 lb/MMBtu  Actual ESP Loading    25.00 lb/MMBtu  Max ESP Loading       25.00 lb/MMBtu</p>	<p>Sorbent Utilization  Material Bal            48.13X  Definition              48.13X    mols SO2 Removed/(mol Na2CO3+ mol CaO)</p>	

(continued)

## APPENDIX D. COMPUTER MODEL OUTPUTS

### TABLE D-1 (continued)

28-Jun-91C O N S D L R & D      GENERIC--250 MW -- Pgh & Coal -- 1.5% S -- Coolside

Capital Costs, \$MM	New	Retro	FacRetrofit
-----			
Sorbent Handling	0.879	1.22	1.071
Lime Preparation	2.330	1.22	2.860
Soda Ash System	0.686	1.22	0.836
FG Handling	3.279	1.34	4.390
Humidifier	1.796	1.34	2.406
Recycle Sys/Injection	0.411	1.22	0.501
Particulate Collection	0.185	1.22	0.226
Reheat	0.000	1.22	0.000
Waste Handling	0.654	1.41	0.924
No Chimney Costs	0.000	1.06	0.000
Misc.	0.613		0.792
-----			
Total Direct	10.833	1.29	13.986
-----			
Field Costs      13.80%	1.495		1.930
Home Office      22.40%	2.427		3.133
Bond,ARI,Tax      1.00%	0.106		0.140
-----			
Contingency      18.00%	14.863		19.188
	2.675		3.454
-----			
Total Plant Investment	17.338		22.642
\$/kw gross	66.836		86.289

	Unit 1	Unit 2	Unit 3	Unit 4	Unit 5	Unit 6	Station
Capacity	262.4	0.0	0.0	0.0	0.0	0.0	262.4
SO2 Removal, %	163.33	0.00	0.00	0.00	0.00	0.00	163.33
Flue Gas, MACFM	879.3	0.0	0.0	0.0	0.0	0.0	879.3
Scrubbed Gas, MACFM	2051.7	0.0	0.0	0.0	0.0	0.0	2051.7
Scrubbed Gas @ Humid Temp,	762.9	0.0	0.0	0.0	0.0	0.0	762.9
Lime, tph	3.40	0.00	0.00	0.00	0.00	0.00	3.40
Solid Waste, tph	6.23	0.00	0.00	0.00	0.00	0.00	6.23
Operating Humidifiers	1	0	0	0	0	0	1
Spare Humidifiers	0	0	0	0	0	0	0
Total Humidifiers	1	0	0	0	0	0	1
Humidifier Length	56.25	0.00	0.00	0.00	0.00	0.00	56.25
Operators per shift	2.33	0	0	0	0	0	2.33
Total Solids in Gas	29.66	0.00	0.00	0.00	0.00	0.00	29.66
Recycle Solids	17.18	0.00	0.00	0.00	0.00	0.00	17.18

#### HUMIDIFIER DESIGN

NO. OPERATING	1
NO. SPARES	0
TOTAL	1
dp, inches H2O	0.025
MACFM/HUMIDIFIER @ OUTLET	763
WIDTH = HEIGHT, FT	27.1
LENGTH, FT	56.2
RESIDENCE TIME, SEC	3
OUTLET VELOCITY, FPS	17.5

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-2

COOLSIDE PROCESS MATERIAL BALANCE\*

28-Jun-91C O N S O L R & D      GENERIC--250 MW -- Pgh & Coal -- 1.5% S -- Coolside

STREAM NO. DESCRIPTION	MATERIAL BALANCE OUTPUT									
	1		2		3		4		5	
	FLUE GAS INPUT		VENT TO STACK		MAKEUP SORBENT		HYDRATOR WATER		HYDRATOR VENT	
	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR
H2O(v)	5.593	100.763	11.648	209.842	---	---	---	---	0.138	2.481
CO	0.004	0.098	0.004	0.098	---	---	---	---	---	---
CO2	10.997	483.982	11.007	484.432	---	---	---	---	---	---
N2	68.802	1927.837	71.457	2002.220	---	---	---	---	---	---
O2	5.350	171.185	6.046	193.534	---	---	---	---	---	---
SO2	0.083	5.306	0.025	1.592	---	---	---	---	---	---
MOISTURE	---	---	0.000	0.002	---	---	0.248	4.468	---	---
CaO	---	---	---	---	0.110	6.184	---	---	---	---
Ca(OH)2	---	---	0.001	0.044	---	---	---	---	---	---
Na2CO3	---	---	0.000	0.000	---	---	---	---	---	---
Na2SO3	---	---	0.000	0.009	---	---	---	---	---	---
Na2SO4	---	---	0.000	0.003	---	---	---	---	---	---
CaSO3	---	---	0.000	0.041	---	---	---	---	---	---
CaSO4	---	---	0.000	0.016	---	---	---	---	---	---
INERTS	---	---	---	0.006	---	0.612	---	---	---	---
C	0.129	1.555	0.001	0.015	---	---	---	---	---	---
FLYASH	---	10.765	---	0.703	---	---	---	---	---	---
TOTAL	90.958	2701.492	100.190	2891.957	0.110	6.795	0.248	4.468	0.138	2.481
TEMP, F	304		149		60		60		212	
PRES, PSI(IN WC)	(8.5)		0		0		60		0	

STREAM NO. DESCRIPTION	MATERIAL BALANCE OUTPUT									
	6		7		8		9		10	
	HUMID FEED		MAKEUP WATER		ADDITIVE MAKEUP		HUMID WATER FEED		WASTES TO DISPOSAL	
	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR
H2O(v)	0.001	0.014	---	---	---	---	---	---	---	---
CO	---	---	---	---	---	---	---	---	---	---
CO2	---	---	---	---	---	---	---	---	---	---
N2	0.054	1.512	---	---	---	---	---	---	---	---
O2	0.014	0.459	---	---	---	---	---	---	---	---
SO2	---	---	---	---	---	---	---	---	---	---
MOISTURE	---	---	6.004	108.172	0.140	2.526	6.004	108.172	0.014	0.248
CaO	---	---	---	---	---	---	---	---	---	---
Ca(OH)2	0.110	8.170	---	---	---	---	---	---	0.062	4.586
Na2CO3	---	---	---	---	0.010	1.082	0.010	1.082	---	---
Na2SO3	---	---	---	---	---	---	---	---	0.008	0.956
Na2SO4	---	---	---	---	---	---	---	---	0.003	0.359
CaSO3	---	---	---	---	---	---	---	---	0.035	4.263
CaSO4	---	---	---	---	---	---	---	---	0.012	1.610
INERTS	---	0.612	---	---	---	---	---	---	---	0.606
C	---	---	---	---	---	---	---	---	0.128	1.540
FLYASH	---	---	---	---	---	---	---	---	---	10.662
TOTAL	0.179	10.767	6.004	108.172	0.150	3.608	6.014	109.255	0.261	24.832
TEMP, F	180		60		115		61		141	
PRES, PSI(IN WC)	10.0		60.0		180.0		180.0		0.0	

\* See Figure D-1 for stream locations.

(continued)

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-2 (continued)

28-Jun-91C O M S O L R & D      GENERIC--250 MW -- Pgh B Coal -- 1.5% S -- Coolside

PRES, PSI(IN WC)	10.0	60.0	180.0	180.0	0.0					
STREAM NO.	11		12		13		14		15	
DESCRIPTION	RECYCLE SOLIDS		COLLECTED SOLIDS		ATOMIZING AIR		SHIELD AIR		HUMID EXIT FLOW	
	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR	M-MOL/HR	M-LB/HR
H2O(v)	0.002	0.027	---	---	---	---	0.014	0.255	11.648	209.842
CO	---	---	---	---	---	---	---	---	0.004	0.098
CO2	---	---	---	---	---	---	---	---	11.007	484.432
N2	0.108	3.025	---	---	1.481	41.490	1.012	28.355	71.457	2002.220
O2	0.029	0.918	---	---	0.394	12.596	0.269	8.608	6.048	193.534
SO2	---	---	---	---	---	---	---	---	0.025	1.592
MOISTURE	0.019	0.348	0.033	0.597	---	---	---	---	0.033	0.599
CaO	---	---	---	---	---	---	---	---	---	---
Ca(OH)2	0.087	6.436	0.149	11.022	---	---	---	---	0.149	11.066
Na2CO3	---	---	---	---	---	---	---	---	---	---
Na2SO3	0.011	1.342	0.018	2.298	---	---	---	---	0.018	2.307
Na2SO4	0.004	0.504	0.004	0.863	---	---	---	---	0.006	0.867
CaSO3	0.050	5.982	0.085	10.245	---	---	---	---	0.086	10.286
CaSO4	0.017	2.260	0.028	3.870	---	---	---	---	0.029	3.885
INERTS	---	0.850	---	1.456	---	---	---	---	---	1.462
C	0.180	2.161	0.308	3.701	---	---	---	---	0.309	3.715
FLYASH	---	14.961	---	25.624	---	---	---	---	---	25.727
TOTAL	0.505	38.813	0.628	59.675	1.874	54.086	1.295	37.219	100.818	2951.632
TEMP, F	180		141		254		60		141	
PRES, PSI(IN WC)	10.0		---		150.0		0.0		(14.50)	

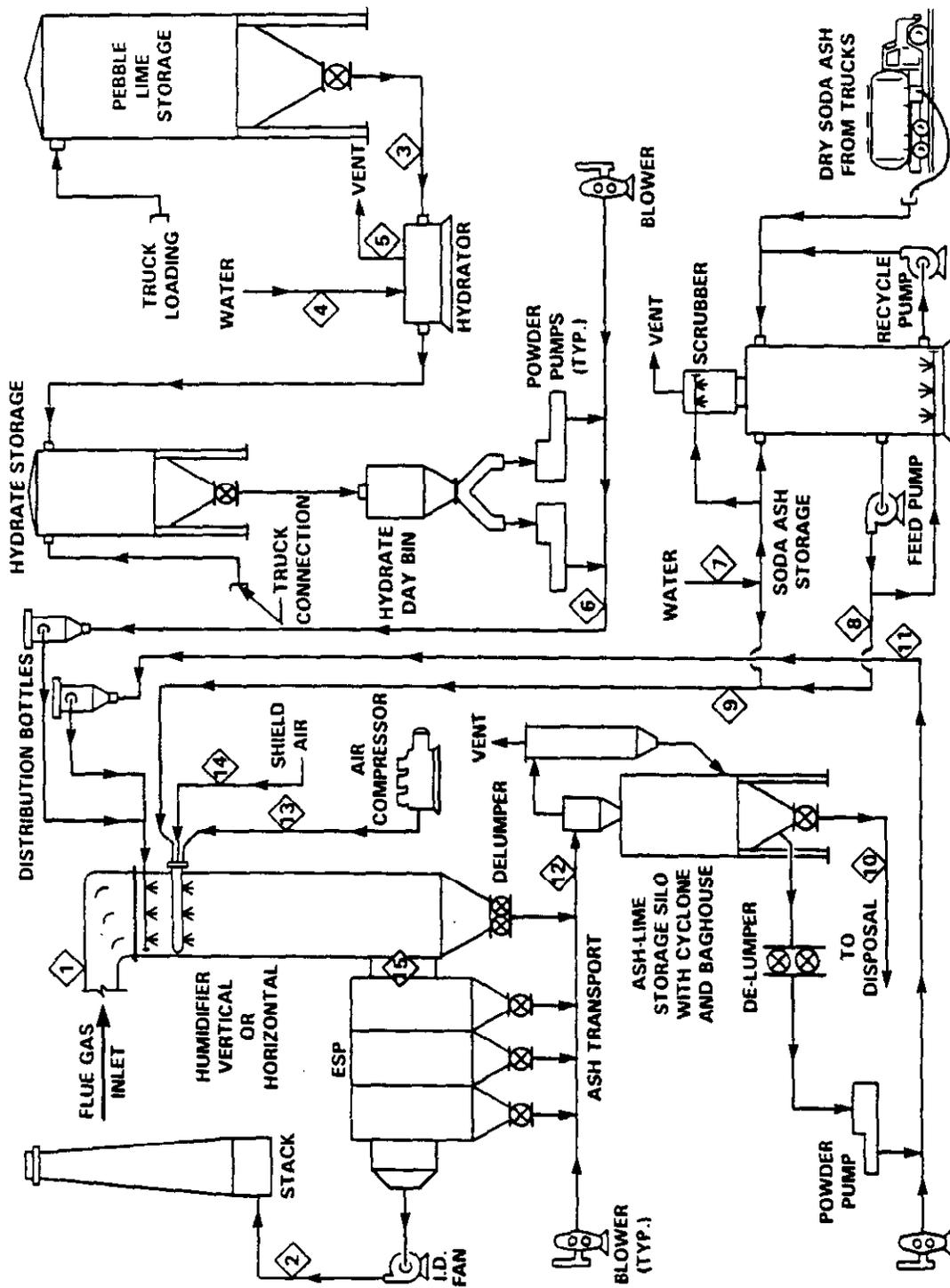


Figure D-1 Coolside process flow diagram with material balance stream identification.

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-3

COOLSIDE PROCESS DIRECT EQUIPMENT COST BASIS

PAGE 1 OF 3

COOLSIDE EQUIPMENT COST ESTIMATE  
\*\*\*\*\*

PROJECT TITLE	GENERIC--250 MW - Coolside Example
PLANT CAPACITY, MW	262.4
HUMIDIFIER SITE, LxWxH (Label)	27.1'x27.1'x56.2'
FEED TYPE	Quicklime (Toggle Quicklime or Hydrate)
NUMBER OF HUMIDIFIERS	1 (Toggle 1 OR 2)
QUICKLINE UNLOAD RATE, MLB/H RATE, TPH	40.77 (0 if by Truck) 20.39
QUICKLINE/M.U.SORBENT, MLB/H RATE, TPH	6.8 3.40
DESIRED HYDRATOR SIZE, TPH	12.99
EQUIV QUICKLINE RATE, TPH	10.00
HYDRATOR RESIDUE RATE, TPH	5.00
HYDRATE TRANSFER RATE, TPH	13.0
TOTAL HYDRATE INJECT, MLB/H " , TPH	8.8 4.39 (Usable Range; 3-20 TPH)
SODA ASH ADD RATE(dry),MLB/H " , TPH	1.1 0.54
VENT FLOW & HUMID OUT, MACFH ESP SCA To Rapper Upgrade	774.7 300
TOTAL ATOMIZATION AIR,MLB/H " , ACFM	54.1 13,070
MAIN AIR COMPRESSOR COMPRESSOR DISCH, PSIG NO OF UNITS IN OPER NO OF SPARE UNITS % CAPACITY/UNIT	150 (Toggle 150 or 180 PSIG) 1 0 100%
WASTER DISPOSAL RATE, MLB/H " , TPH	24.8 12.42

(continued)

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-3 (continued)

COOLSIDE EQUIPMENT COST ESTIMATE (CONT)

```

*****
PROJECT TITLE          GENERIC--250 MW - Coolside Example
HUMIDIFIER SITE, LxWxH 27.1'x27.1'x56.2'

RECYCLE SOLIDS RATE, MLB/H    34.8
      "           , TPH      17.62

TOTAL COLLECT SOLIDS, MLB/H   59.7
      "           , TPH      29.84

ESP SCALPER             No      (Toggle Yes or No)

ASH-HYDRATE STORAGE SILO  Yes    (Toggle Yes or No)
    
```

DUCTWORK & ACCESSORIES - SURFACE AREAS (SQFT)

```

*****
HUMIDIFIER (EACH)

I/O HUMID WIDTH, FT    15.9
I/O DUCT WIDTH, FT    11.3

HUMIDIFIER WIDTH, FT  27.1
      "   LENGTH, FT  56.2

INLET/OUTLET DUCT TOTAL LENGTH, FT    160.0

DUCTS TO/FROM EA HUMIDIFIER    2 (1 (normal) or 2)

BYPASS DUCTWORK AND ACCESSORIES, TOTAL FT2    0

NUMBER OF DAMPERS, TOTAL DAMPER AREA    4    127.1
    
```

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-3 (continued)

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### COOLSIDE EQUIPMENT COST ESTIMATE (CONT)

PROJECT TITLE	GENERIC--250 MW - Coolside Example			
HUMIDIFIER SITE, LxWxH	27.1'x27.1'x56.2'			
	QUICKLINE	TOTAL	ASH	PUGMILL
	UNLOADING	COLLECTED	RECYCLE	
MISCELLANEOUS SYSTEMS	BY RAIL	SOLIDS	(/HUMID)	SOLIDS
-----	-----	-----	-----	-----
NUMBER OF UNITS OPERATING	1	1	1	1
NUMBER OF UNITS SPARE	0	0	0	0
	QUICKLINE	RECYCLE		HYDRATOR
CONCRETE STORAGE SILOS	FEED	ASH/LIME		SPARES
-----	-----	-----		-----
BULK DENSITY, LB/CUFT	55	35		1
STORAGE, DAYS	15	4		
STORAGE, HOURS	0	0		
NUMBER OF SILOS	1	1		
	HYDRATOR	HYDRATOR	TRANSPORT	STORAGE
STEEL TANKS & BINS	FEED BIN	RESIDUE	FEED BIN	DAY BIN
-----	-----	-----	-----	-----
BULK DENSITY, LB/CUFT	55	30	30	30
NUMBER OPERATING	2	2	2	1
NUMBER SPARES	0	0	0	0
STORAGE, HOURS	1	0.5	1	24

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-4

COOLSIDE PROCESS DIRECT EQUIPMENT COST SUMMARY

PAGE 1 OF 2

SUMMARY OUTPUT  
 \*\*\*\*\*  
 GENERIC--250 MW - Coolside Example  
 27.1'x27.1'x56.2'

WATER, GPM	QUANTITY	SIZE, EACH	PROCESS EQUIPMENT	CONTD HP	OPER HP	BULK FACTOR	DIRECT COST, \$
*****	*****	*****	*****	*****	*****	*****	*****
--	1	PK	RAIL UNLOADING STATION	125	21	2.75	\$370,608
--	1	44 MFT3	CONCRETE FEED STORAGE SILO	5	5	1.00	\$230,526
--	1	10.0 TPH	QUICKLINE PNEUMATIC TRANS PK	169	96	2.40	\$228,000
--	2	275 FT3	HYDRATOR FEED BIN(S)	10	10	2.40	\$65,268
80	2	10.0 TPH	LIME HYDRATOR PACKAGE	150	25	3.00	\$1,200,000
--	2	97 FT3	HYDRATOR RESIDUE BIN(S)	10	10	2.40	\$38,641
--	1	5.0 TPH	RESIDUE TRANSPORT PACKAGE	40	40	2.40	\$40,800
--	2	171 FT3	HYDRATE TRANSPORT FEED BIN	10	10	2.40	\$51,296
--	1	13.0 TPH	HYDRATE PNEUMATIC TRANS PK	214	161	2.30	\$409,400
--	1	8,184 FT3	HYDRATE STORAGE DAY BIN(S)	5	5	2.40	\$179,771
--	1	4.4 TPH	HYDRATE INJECTION PACKAGE	108	61	2.00	\$203,680
--	1	EA	HYDRATE DISTRIBUTION BOTTLES	--	--	2.00	\$9,274
60	1	0.54 TPH	SODA ASH STORAGE/FEED SYSTEM	35	35	2.50	\$647,084
--	1	EA	ATOMIZ LIQUID/AIR HARDWARE	40	20	2.00	\$417,800
--	271	EA	ATOMIZATION NOZZLES	--	--	1.00	\$108,400
425	1	15,000 ICFM	AIR COMPRESSORS (150 PSIG)	3,500	3,050	2.40	\$998,400
--	1	565 GPM	TOTAL WATER SUPPLY PREP	100	50	2.40	\$59,808
--	1	39,868 FT2	DUCTWORK & TURNING VANES HUMIDIFIER	30	30	2.07	\$2,036,663
--	1	0 FT2	BYPASS MODIFICATION PK	--	--	2.07	\$0
--	160	7,217 FT2	INLET/OUTLET PK	--	--	2.07	\$442,233
--	4	127.1 FT2	DAMPERS	40	0	2.07	\$614,053

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-4 (continued)

PAGE 2 OF 2

SUMMARY OUTPUT  
\*\*\*\*\*  
GENERIC--250 MW - Coal-side Example

27.1'x27.1'x56.2'

WATER, GPM	QUANTITY	SIZE, EACH	PROCESS EQUIPMENT	CONTD HP	OPER HP	BULK FACTOR	DIRECT COST, \$
*****	*****	*****	*****	*****	*****	*****	*****
--	0	EA	ESP SCALPER	--	--	--	--
--	1	EA	DUCT SOOTBLOWERS & RAPPERS	10	10	2.40	\$50,400
--	1	60 MFT3	ASH-HYDRATE STORAGE SILO	5	5	1.00	\$302,915
--	1	74.5 TPH	PUG MILL	100	17	2.70	\$316,126
--	1	17.4 TPH	ASH RECYCLE INJECTION EQUIP	125	125	2.75	\$84,608
--	1	EA	WASTE WATER TREATMENT	10	10	1.25	\$45,000
--	1	EA	AIR FOIL BRUSH CLEANERS	20	20	1.00	\$60,000
--	1	119.4 TPH	ESP ASH TRANSFER UPGRADE	450	113	2.75	\$255,833
--	1	PK	ESP RAPPER UPGRADE PACKAGE	--	--	1.00	\$174,587
TOTALS				5,311	3,926		\$9,641,156

"ACTUAL" OPERATING HORSEPOWER 3,337

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-5

COOLSIDE PROCESS EQUIPMENT LISTING

Quicklime Test

Item Number	Equipment Number	Qty	Description	Total Power/ Services
1	UP-101	1	Rail Quicklime Unloading Package Package Includes: 1 Blower System with 0 Spare Blowers and Rail Spur.	21 HP
2	SS-102	1	Quicklime Storage Silo(s) 15 Day Supply @ Bulk Density of 55 lb/ft <sup>3</sup> , Dia. 28 Ft., Ht. 71 Ft., Concrete Construction.	5 HP
3	QT-103	1	Quicklime Transport to Hydrator Package Includes: 2-Blowers, 2-Bin Vents, 2-Hopper Activators, 2-Feeders, 350 Ft. Pipe and Valves.	169 HP
4	BN-104	2	Hydrator Feed Bin, 2 Bins Operating, Providing 1.0 Hours of Storage @ Bulk Density of 55 lb/ft <sup>3</sup> , and 0 Spare.	10 HP
5	HP-105	2	Hydrator Package and 1 Spare  Package Includes: Feeder, Mixers, Scrubber, and Controls.	150 HP
6	BN-106	2	Hydrator Residue Receiving Bin 2 Bins Operating, Providing 0.5 Hours of Storage @ Bulk Density of 30 lb/ft <sup>3</sup> , and 0 Spare.	10 HP
7	RT-107	1	Hydrator-Residue Transport Package Package Includes: 1-Blower, 2-Feeders, Pipe and Valves.	40 HP
8	BN-108	2	Hydrate Feed Bin 2 Bins Operating, Providing 1.0 Hours of Storage @ Bulk Density of 30 lb/ft <sup>3</sup> , and 0 Spare.	10 HP

(continued)

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-5 (continued)

Quicklime Test

Item Number	Equipment Number	Qty	Description	Total Power/ Services
9	HT-109	1	Hydrate Transport Package Package Includes: 3-Blowers, 1-Fluidizing Blower, 1-Bin Vents, 2-P. P., Pipe and Valves.	214 HP
10	BN-110	1	Day Bin for Hydrate Storage 1 Bin Operating, Providing 24 Hour Supply @ Bulk Density of 30 lb/ft <sup>3</sup> , and 0 Spare.	5 HP
11	HI-111	1	Hydrate Injection Package Package Includes: 3-Blowers, 1-Fluidizing Blower, 2-Injection Streams, 3-P. P., Pipe and Valves. (Volumetric Injection)	108 HP
12	DB-112	1	Distribution Bottles -- Hydrate	--
13	SA-113	1	Soda Ash Feed/Storage Package Package Includes: Feed Pump, Recycle Pump, Tank, and Scrubber.	35 HP 60 GPM
14	LA-114	1	Atomization Liquid/Air Feed Package Package Includes: 2-Inline Mixers, 2-Pumps, Filters, Airfall, Headers, Pipe, Valves and Instrumentation.	40 HP
15	AN-115	271	Atomization Nozzles	--
16	CA-116	1	Air Compressors 150 PSIG Discharge Pressure with 1 Units Operating @ 100 % Capacity and 0 Spare.	3500 HP 425 GPM
17	WP-117	1	Water Supply Package Supplies 565 GPM Package Includes: 2-Pumps, 2-Primary Filters.	100 HP

(continued)

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-5 (continued)

Quicklime Test

Item Number	Equipment Number	Qty	Description	Total Power/ Services
18	MD-118	1	Humidifier Duct 23162 Square Feet of Ducting and 16706 Square Feet of Vanes (Total). Includes: Ash Feeders.	30 HP
19	SD-119	1	Bypass Duct 0 Square Feet (Total).	--
20	ID-120	1	Inlet/Outlet Ducting 7217 Square Feet (Total).	--
21	DA-121	4	Dampers 127 Square Feet (Each), Motorized.	40 HP
22	SC-122		ESP Scalper with Ash Feeders.	0 HP
23	SB-123	1	Duct Sootblower and Rapper Package (Uses High Pressure Air from Existing Plant Equipment.) Package Includes: Pipe and Valves.	10 HP
24	SS-124	1	Ash-Hydrate Silo 4 Day Storage @ Bulk Density of 35 lb/ft <sup>3</sup> Dia. 33 Ft., Ht. 82 Ft., Concrete Construction. Package Includes: Cyclone, Baghouse and Truck Loadout.	5 HP
25	PM-125	1	Pugmill 1 Units Operating and 0 Spare.	100 HP
26	AI-126	1	Ash Recycle Injection Package Each Package Includes: 1 Operating Blower with 0.0 Spare Blower, 1-P. P., Distribution Bottle, Delumper, Injection Hose and Pipe.	125 HP

(continued)

APPENDIX D. COMPUTER MODEL OUTPUTS

TABLE D-5 (continued)

Quicklime Test

Item Number	Equipment Number	Qty	Description	Total Power/ Services
27	WT-127	1	Waste Water Treatment For Floor Drains and Area Run-off.	10 HP
28	BC-128	1	Duct Cleaner Brushes for Cleaning Nozzles and Lances/Airfoils.	20 HP
29	HE-129	1	Flue Gas Reheater Package (NOT INCLUDED) Low pressure Steam Operation.	--
30	ST-130	1	Stack Relining (NOT INCLUDED).	--
31	ES-131	1	ESP Modification Package (NOT INCLUDED) Package Includes: Hopper Insulation and Heaters, Fluidization Air Piping.	--
32	AC-132	1	ESP Ash Transport Upgrade Upgrade Includes: 1 Operating Blower Systems with 0 Spare Blowers.	450 HP
33	AC-133	1	ESP Rapper Upgrade Package	--

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6

### LIMESTONE FORCED OXIDATION FGD ECONOMICS SUMMARY

CONSOL Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh B Coal -- 1.5% S Fuel--Limestone Forced Oxidation    28-Jun-91

Unit: GENERIC--250 MW                      Coal: Pgh B Coal -- 1.5%  
Forced Ox.

#### ASSUMPTIONS

Unit	GENERIC--250 MW	No. Units	1		
Fuel	Pgh B Coal -- 1.5%MW, Net		249.6	246.4	
MW, Gross	262	Aux power	4.89%	6.12%	
Net Heat Rate	9,506	9,631 Turb H R	7,861	Heatlosses	
% Excess Air	40.00	Spare Piv	100%	(C)	0.913% Fuel Btu
Air Temp	60	Spare Abs.	0%	CO	0.018% "
F G Exit Temp	304.0	Reheat	0 deg	Manuf	1.500% "
SO2 Removal, %	95.00	Cap Fact	65.00%	Rad	0.187% "
Ca/S Ratio	1.05	Base Year	1990		
Lb H2O/lb bdeir	0.0069	Max Abs D	58 ft		
flyash/ash	80.00%	Act Abs D	38.28 ft		
Plant Loc fac	1.06	Unloading	1 0=Truck 1=Rail		
F/O = 1	1	Retrofit	1 0=New 1=Retrofit		
		ID Fan Suc	-14.5 in wg		

#### Compositions, weight % dry basis

Coal		Limestone, wt%		Lime, wt%	
H	4.93	CaCO3	95.00	CaO	95.50
C	79.89	MgCO3	0.00	MgO	3.00
H + Cl	1.52	Inerts	5.00	Inerts	1.50
D	4.03				
S	1.59				
Ash	8.04				
NHV	14,180	14,427 Dulong			
Stoic Air	10.71				
Moist	5.50				

#### RESULTS

Efficiency	86.94%	SO2 Removal	95.00%
Net Heat Rate	9,506	SO2 Emissions	0.11 lb/MMBtu
O2,dry	6.28%		755 ton/yr
FlueGas,MSCFM	574.6		0.13 tph
MACFM	886.1	-11.5 "H2O @ Air Heater	
	893.0	-14.5 "H2O @ I D Fan Suction	
Particulate loading	5.06 lb/MMBtu		
	1.581 g/ecf		
Flue Gas Composition, mol % or ppm		Particulate, weight %	
	Dry	Moist	
CO	41	39	(C) 10.36%
CO2	12.90%	12.11%	Ash+Inert 89.64%
H2O	0.00%	6.16%	
N2	80.72%	75.75%	
O2	6.28%	5.89%	
SO2	972	912	
Mol Wt.	30.37	29.61	

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOL Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5X \$ Fuel--Limestone Forced Oxidation    28-Jun-91

Unit: GENERIC--250 MW                      Coal: Pgh 8 Coal -- 1.5X  
Forced Ox.

				Levelized					
Operating Costs	Units	\$/Unit	Units/yr	Factor	\$/MWh/yr	mill/kWh	\$/t Coal	\$/t SO <sub>2</sub>	R
<b>Major Consumables</b>									
Limestone	ton	15.00	24,782	1.000	0.372	0.26	0.74	26	
Fixation Lime	ton	60.00	0	1.000	0.000	0.00	0.00	0	
Import Flyash	ton	5.00	0	1.000	0.000	0.00	0.00	0	
Water	Mgal	0.650	87,537	1.000	0.057	0.04	0.11	4	
Steam	MMBtu	3.50	0	1.000	0.000	0.00	0.00	0	
Waste Disposal (dry)	ton	7.00	40,438	1.000	0.283	0.20	0.56	20	
Flyash Credit	ton	(7.00)	0	1.000	0.000	0.00	0.00	0	
Power	MWh	29.00	18,383	1.000	0.533	0.38	1.06	37	
<b>Total Variable</b>					1.245	0.88	2.47	87	
<b>Fixed O&amp;M</b>									
Operating Labor		22.92	33,288	1.000	0.763	0.54	1.51	53	
Maintenance				1.000	1.114	0.78	2.21	78	
Admin overhead				1.000	0.363	0.26	0.72	25	
<b>Total Fixed</b>					2.239	1.58	4.44	156	
<b>Total O&amp;M</b>					3.484	2.45	6.91	243	
<b>Capital Costs</b>									
Total Plant Inv		44.053	0.1180	1.000	5.199	3.66	10.31	362	
Preproduction		1.227	0.1180	1.000	0.145	0.10	0.29	10	
Working Capital		0.268	0.1180	1.000	0.032	0.02	0.06	2	
Int. During Const. (IDC)		2.742							
<b>Total Capital</b>					5.375	3.78	10.66	375	
<b>Total Cost</b>					8.859	6.23	17.58	617	

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOL Coal Quality Model - FGD Module GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5% S Fuel--Limestone Forced Oxidation 28-Jun-91

Unit: GENERIC--250 MW      Coal: Pgh 8 Coal -- 1.5%  
Forced Ox.

Plant	GENERIC--250 MW					Reagent Preparation	
Capacity	249.58 MW net ex FGD	246.35 MW net incl. FGD			*****		
Coal	Pgh 8 Coal -- 1.5% S					NA Max Silo Diam., ft	
%S as received	1.50 %				NA Max Silo Ht., ft		
HHV as rec.	13,400 Btu/lb				2 Total Pulverizers		
feed rate, base	88.53 ton/hr				100 % per Pulverizer		
Capacity Factor	65%	Heat Rate:	9,506 ex FGD	9,631 incl FGD	Fan		
Retrofit Diff.	MEDIUM Site Access	MEDIUM Undernd Obst			*****		
Ca/S mole ratio	1.05				2 NUMBER/unit		
Wet bulb temp	121.7 F	Power Consumption			5.5 Pres Drop, in. H2O		
Water added	0.0470 lb/lb fg	-----			Absorber		
	243.4 gpm	Reagent Prep & Handling			*****		
	43 MSCFM	Booster Fan			94 kW		
Absorber Flue Gas	617 MSCFM	Pumps			1012 kW	0.00 Velocity, fps co-cur	
	691 MACFM	Oxidation			1822 kW	10.00 Velocity, fps contr-cur	
Bypass Flow	0.0%	Waste Disposal			113 kW	100.93 L/G, gal/MSCF	
	0 MACFM	Misc.			124 kW	1.05 Ca/S Mol Ratio	
Mix T <sub>w</sub>	121.7 deg F				63 kW	95.0 Design Removal	
FG Cp, estimated	0.26	Total			3,229 kW	1.5 O2/SO2 mol ratio	
Reheat Req	0 deg F	-----			0.67 Inlet/Spray Tur		
	0.0 MMBtu/hr Steam				0.00 CaSO3/CaSO4 Mol Ratio		
No. of Absorbers	1 operating	0 spare	1 total		Duct Runs		
					*****		
Limestone	4.35 tph				0 By-Pass (Yes=1)		
					200 To Absorbers		
					75 From Absorbers		
SO2 Removed	2.52 tph	Thickener	1	61 ft dia	Thickeners		
	14,352 ton/yr	Filters	2	163 sq ft	*****		
	95%	1 operating			1 spare	15 sq ft/dtpd	
SO2 emission	0.13 tph	Flyash/DFC	0.87		10.00 % feed solids		
	755 ton/yr	Flyash req.	0.00 tph		30.00 % underflow solids		
	0.112 lb/MMBtu	Import Flyash	0.00 tph		200 max diameter, ft		
Solid Waste	7.10 tph dry filter cake	Fixation Lime	0.00 tph		0 spare		
	7.10 tph dry waste	Wet Solid Waste	8.88 tph		Filters		
Power	3,229 kW	Chlorine Content	56,225 ppm		*****		
Water	256.2 gpm				1 = Filter 2=Centrifuge		
	Unit 1	Unit 2	Unit 3	Unit 4	Unit 5	Unit 6	Station
Capacity	262.4	0.0	0.0	0.0	0.0	0.0	262.4
SO2 Removal, %	95.00	0.00	0.00	0.00	0.00	0.00	95.00
Flue Gas, MACFM	886.1	0.0	0.0	0.0	0.0	0.0	886.1
Scrubbed Gas, MACFM	886.1	0.0	0.0	0.0	0.0	0.0	886.1
Absorber Flow, MACFM	690.6	0.0	0.0	0.0	0.0	0.0	690.6
Limestone, tph	4.35	0.00	0.00	0.00	0.00	0.00	4.35
Solid Waste, tph	7.10	0.00	0.00	0.00	0.00	0.00	7.10
Operating Absorbers	1	0	0	0	0	0	1
Spare Absorbers	0	0	0	0	0	0	0
Total Absorbers	1	0	0	0	0	0	1
Absorber Dia., ft	38.28	0.00	0.00	0.00	0.00	0.00	38.28
Operators per shift	3.8	0	0	0	0	0	3.8
	*****						
	150 lb/hr/sq ft rate						
	2 Shift/day						
	50 % Spare						
	80.00 % Cake Solids						
	700 max area, sq ft						
	Fixation						
	NA Flyash/DFC req.						
	NA Lime/DFC						
	0 Chimney						
	1 *****						
	1 0=No Chimney Cost						
	1 1=New 0=Reline						

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOI Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5% \$ Fuel--Limestone Forced Oxidation    28-Jun-91

Unit: GENERIC--250 MW            Coal: Pgh 8 Coal -- 1.5%  
Forced Ox.

Capital Costs,\$MM	Unit 1	Unit 2	Unit 3	Unit 4	Unit 5	Unit 6	Station	Retro	Fac
Limestone Handling	NA	NA	NA	NA	NA	NA	1.600	1.22	1.951
Limestone Prep	NA	NA	NA	NA	NA	NA	2.524	1.22	3.076
FG Handling	3.609	0.000	0.000	0.000	0.000	0.000	3.609	1.34	4.833
SO2 Removal	7.387	0.000	0.000	0.000	0.000	0.000	7.387	1.34	9.891
Reheat	0.000	0.000	0.000	0.000	0.000	0.000	0.000	1.22	0.000
Waste Handling	NA	NA	NA	NA	NA	NA	1.917	1.41	2.709
Chimney	3.010	0.000	0.000	0.000	0.000	0.000	3.010	1.06	3.191
Misc.	NA	NA	NA	NA	NA	NA	1.203		1.539
<b>Total Direct</b>							<b>21.250</b>	<b>1.28</b>	<b>27.191</b>
Field Costs	13.80%						2.933		3.752
Home Office	22.40%						4.760		6.091
Bond,ARI,Tax	1.10%						0.234		0.299
							<b>29.176</b>		<b>37.333</b>
Contingency	18.00%						5.252		6.720
<b>Total Plant Investment</b>							<b>34.428</b>		<b>44.053</b>
\$/kw gross							131.205		167.883

Retrofit Factors	Site Access	Ungrd Obstruct	Soil	TOTAL
	MEDIUM	MEDIUM		
Limestone Handling	1.15	1.06	1.00	1.22
Limestone Prep	1.15	1.06	1.00	1.22
FG Handling	1.30	1.03	1.00	1.34
SO2 Removal	1.30	1.03	1.00	1.34
Reheat	1.15	1.06	1.00	1.22
Waste Handling	1.24	1.14	1.00	1.41
Stack	1.06	1.00	1.00	1.06

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOL Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5% S Fuel--Limestone Forced Oxidation    28-Jun-91

Unit: GENERIC--250 MW                      Coal: Pgh 8 Coal -- 1.5%  
Forced Ox.

```

=====
SO2 ABSORPTION SECTION
=====
ABSORBERS
=====
NO. OPERATING           1         0         0         0         0         0
NO. SPARES              0         0         0         0         0         0
TOTAL                   1         0         0         0         0         0
dP, inches H2O          3.5       0.0       0.0       0.0       0.0       0.0
MACFM/ABSORBER         691       0         0         0         0         0
L/G, gal/1000 SCF      100.9     0.0       0.0       0.0       0.0       0.0
GPM WATER FLOW         69,700    0         0         0         0         0
NO SPRAY HEADERS       4         0         0         0         0         0
DIAMETER, ft           38.3     0.0       0.0       0.0       0.0       0.0
SPRAY TOWER HEIGHT, ft 53.5     0.0       0.0       0.0       0.0       0.0
ABSORBER SUMP
VOLUME, CU FT          93,182    0         0         0         0         0
DIA, FT                49.1     0.0       0.0       0.0       0.0       0.0
HT, FT                 49.1     0.0       0.0       0.0       0.0       0.0
RESIDENCE TIME, MIN    10        0         0         0         0         0
TOTAL SURFACE, SQ FT   18,379    0         0         0         0         0
COST/ABSORBER, $1000   $1,868    $0        $0        $0        $0        $0
TOTAL ABSORBER, $1000 $1,868    $0        $0        $0        $0        $0

PUMPS
=====
OPER PUMPS/ABS         4         0         0         0         0         0
TOTAL PUMPS/ABS        5         0         0         0         0         0
PUMP HEAD, FT          100       0         0         0         0         0
GPM/PUMP               17,425    0         0         0         0         0
HP/PUMP, DESIGN        912       0         0         0         0         0
HP/PUMP, OPERATING     615       0         0         0         0         0
TOTAL OPERATING HP     2,459     0         0         0         0         0
COST/PUMP, $1000       $99       $0        $0        $0        $0        $0
TOTAL PUMP COSTS       $497     $0        $0        $0        $0        $0

OXIDATION AIR COMPRESSORS
=====
NO OPER                1         0         0         0         0         0
TOTAL                  2         1         1         1         1         1
SCFM/COMP              3,350     0         0         0         0         0
HP/COMP                152       0         0         0         0         0
TOTAL OPERATING HP     152       0         0         0         0         0
COMPRESSOR COST, $1000 $74       $0        $0        $0        $0        $0
TOTAL COMPRESS COST, $1000 $148     $0        $0        $0        $0        $0

TOTAL COSTS
=====
EQUIPMENT, $1000       $2,512    $0        $0        $0        $0        $0
TOTAL COST, $1000     $7,387    $0        $0        $0        $0        $0
=====

```

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOL Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5% S Fuel--Limestone Forced Oxidation    28-Jun-91

		Unit: GENERIC--250 MW Forced Ox.		Coal: Pgh 8 Coal -- 1.5%			
=====							
FLUE GAS HANDLING SECTION							
=====							
FANS (2 FANS/UNIT)							
=====							
ACFM/FAN	DESIGN	487	0	0	0	0	0
	PERFORMANCE	443	0	0	0	0	0
DELTA P, IN H2O	DESIGN	6.6	0.0	0.0	0.0	0.0	0.0
	PERFORMANCE	5.5	0.0	0.0	0.0	0.0	0.0
HP/FAN	DESIGN	871	0	0	0	0	0
	PERFORMANCE	683	0	0	0	0	0
COST/FAN, \$1000		\$128	\$0	\$0	\$0	\$0	\$0
TOTAL FAN COST, \$1000		\$257	\$0	\$0	\$0	\$0	\$0
DUCTWORK							
=====							
DUCT WIDTH	FAN	7.6	0.0	0.0	0.0	0.0	0.0
	INLET	15.2	0.0	0.0	0.0	0.0	0.0
	BY-PASS	0.0	0.0	0.0	0.0	0.0	0.0
	ABS IN	25.6	0.0	0.0	0.0	0.0	0.0
	ABS OUT	15.2	0.0	0.0	0.0	0.0	0.0
	OUTLET	15.2	0.0	0.0	0.0	0.0	0.0
DUCT HEIGHT	FAN	19.5	0.0	0.0	0.0	0.0	0.0
	INLET	19.5	0.0	0.0	0.0	0.0	0.0
	BY-PASS	0.0	0.0	0.0	0.0	0.0	0.0
	ABS IN	11.5	0.0	0.0	0.0	0.0	0.0
	ABS OUT	15.2	0.0	0.0	0.0	0.0	0.0
	OUTLET	15.2	0.0	0.0	0.0	0.0	0.0
DUCT LENGTH	FAN	60.0	0.0	0.0	0.0	0.0	0.0
	INLET	200.0	0.0	0.0	0.0	0.0	0.0
	BY-PASS	0.0	0.0	0.0	0.0	0.0	0.0
	ABS IN	17.4	0.0	0.0	0.0	0.0	0.0
	ABS OUT	36.6	0.0	0.0	0.0	0.0	0.0
	OUTLET	75.0	0.0	0.0	0.0	0.0	0.0
SQ FT SURFACE	FAN	3,247	0	0	0	0	0
	INLET	13,856	0	0	0	0	0
	BY-PASS	0	0	0	0	0	0
	ABS IN	1,295	0	0	0	0	0
	ABS OUT	2,219	0	0	0	0	0
	OUTLET	4,552	0	0	0	0	0
	CHIMNEY TRN	1,779	0	0	0	0	0
DUCT COST	FAN	\$103	\$0	\$0	\$0	\$0	\$0
	INLET	\$438	\$0	\$0	\$0	\$0	\$0
	BY-PASS	\$0	\$0	\$0	\$0	\$0	\$0
	ABS IN	\$41	\$0	\$0	\$0	\$0	\$0
	ABS OUT	\$95	\$0	\$0	\$0	\$0	\$0
	OUTLET	\$194	\$0	\$0	\$0	\$0	\$0
	CHIMNEY TRN	\$76	\$0	\$0	\$0	\$0	\$0
DAMPERS	FAN	\$171	\$0	\$0	\$0	\$0	\$0
	BY-PASS	\$0	\$0	\$0	\$0	\$0	\$0
	ABS IN	\$169	\$0	\$0	\$0	\$0	\$0
	ABS OUT	\$209	\$0	\$0	\$0	\$0	\$0
TOTAL EQUIPMENT		\$1,752	\$0	\$0	\$0	\$0	\$0
TOTAL SECTION COST		\$3,609	\$0	\$0	\$0	\$0	\$0

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOL Coal Quality Model - FGD Module GENERIC--250 MW Plant--Pgh B Coal -- 1.5% S Fuel--Limestone Forced Oxidation 28-Jun-91

Unit: GENERIC--250 MW Coal: Pgh B Coal -- 1.5%  
Forced Ox.

WASTE DISPOSAL SECTION		page 1	
=====			
THICKENER SYSTEM		FILTER SYSTEM	
=====			
MAXIMUM DIAMETER FT	200	NO. OPER	1
NO. OPER	1	NO. SPARE	1
NO. SPARE	0	CLOTH AREA, SQ FT	163
ACTUAL DIAMETER FT	61	CLOTH AREA/FILTER, SQ FT	163
HP/THICKENER	25	DESIGN HP	75
TOTAL OPER HP	25	OPER HP	75
COST/THICKENER	\$207,965	\$/FILTER	\$91,200
\$ TOTAL	\$207,965	\$ TOTAL	\$182,400
=====			
FLOCCULANT		TANKS	
=====			
NO. OPER	1	FEED	RETURN
HP/SYSTEM	10	NO. TANKS	1
TOTAL OPER HP	10	RESIDENCE, HRS EA	12.00
COST/SYSTEM	\$16,548	VOLUME, CU FT	9,560
\$ TOTAL	\$16,548	DIAMETER, FT	27.6
=====			
NO. OPER	1	HEIGHT, FT	16.0
RESIDENCE TIME, min	60	POLY LINING(1=Y,0=N)	1
VOLUME, cu ft	2,563	\$/TANK & AGITATOR	\$81,017
DIAMETER, ft	20.2	\$ TOTAL	\$81,017
HEIGHT, ft	8.0	=====	
POLY LINING(1=Y,0=N)	1	PUMPS	
\$/TANK	\$40,183	=====	
\$ TOTAL	\$40,183	FEED	FILTER
=====			
NO. OPER	1	FEED	RETURN
RESIDENCE TIME, min	60	NO. OPER	1
VOLUME, cu ft	2,563	NO. SPARE	1
DIAMETER, ft	20.2	GPM/PUMP	78
HEIGHT, ft	8.0	FT. HEAD	50
POLY LINING(1=Y,0=N)	1	DESIGN HP	6.00
\$/TANK	\$40,183	HP OPER	5.22
\$ TOTAL	\$40,183	\$/PUMP	\$4,930
=====			
NO. OPER	1	\$ TOTAL	\$9,859
RESIDENCE TIME, min	60	=====	
VOLUME, cu ft	2,563	CONVEYORS	
DIAMETER, ft	20.2	=====	
HEIGHT, ft	8.0	FEED	FILTER CAKE
POLY LINING(1=Y,0=N)	1	NO. OPER	1
\$/TANK	\$40,183	NO. SPARE	1
\$ TOTAL	\$40,183	TON/HR	12
=====			
NO. OPER	1	FT/MIN	120
RESIDENCE TIME, min	60	WIDTH, IN	12
VOLUME, cu ft	2,563	LENGTH, FT	100
DIAMETER, ft	20.2	HP/CONVEYOR	1.6
HEIGHT, ft	8.0	HP	1.6
POLY LINING(1=Y,0=N)	1	\$/CONVEYOR	\$40,009
\$/TANK	\$40,183	\$ TOTAL	\$80,018
\$ TOTAL	\$40,183	=====	
=====			
EQUIPMENT TOTAL	\$291,802	EQUIPMENT TOTAL	\$436,848
BULKS	\$195,156	BULKS	\$387,470
LABOR	\$331,719	LABOR	\$273,577
TOTAL	\$818,678	TOTAL	\$1,097,895
=====			

(continued)

# APPENDIX D. COMPUTER MODEL OUTPUTS

## TABLE D-6 (continued)

CONSOI Coal Quality Model - FGD Module    GENERIC--250 MW Plant--Pgh 8 Coal -- 1.5% S Fuel--Limestone forced Oxidation    28-Jun-91

Unit: GENERIC--250 MW                      Coal: Pgh 8 Coal -- 1.5%  
Forced Ox.

=====				page 2	
WASTE DISPOSAL SECTION					
=====					
WASTE MIXING AND CONVEYING SYSTEM					
=====					
BINS				PUGMILL	
=====				=====	
FLY ASH	NO. BINS	0		NO. OPER	0
	CAPACITY, DAYS	3		NO. SPARE	0
	VOLUME, CU FT	0		HP/MIXER OPER	0
	DIAMETER, FT	0.0		HP DESIGN	0
	HEIGHT, FT	0.0		\$/MIXER	\$0
	\$/BIN	\$0		\$ TOTAL	\$0
	\$ TOTAL	\$0			
FIXATIVE	NO. BINS	0		WASTE CONVEYOR	
	CAPACITY, DAYS	3		=====	
	VOLUME, CU FT	0		NO. OPER	0
	DIAMETER, FT	0.0		NO. SPARE	0
	HEIGHT, FT	0.0		TON/HR	0
	\$/BIN	\$0		FT/MIN	120
	\$ TOTAL	\$0		WIDTH, IN	12
				LENGTH, FT	100
CONVEYORS				HP/CONVEYOR	1.5
=====				HP	0.0
	FLYASH	FIXATIVE		\$/CONVEYOR	\$33,572
NO. OPER	0	0		\$ TOTAL	\$0
NO. SPARE	0	0		=====	
TON/HR	0	0		EQUIPMENT TOTAL	\$0
FT/MIN	120	120		BULKS	\$0
WIDTH, IN	12	12		LABOR	\$0
LENGTH, FT	100	100		TOTAL	\$0
HP/CONVEYO	1.5	1.5		=====	
HP	0.0	0.0		TOTAL SECTION COST	
\$/CONVEYOR	\$33,572	\$33,572		=====	
\$ TOTAL	\$0	\$0		THICKENER	\$818,678
WEIGH BELT				FILTER	\$1,097,895
=====				OTHER	\$0
	FLYASH	FIXATIVE	FGD WASTE	-----	
NO. OPER	0	0	0	TOTAL SECTION COST	\$1,916,573
NO. SPARE	0	0	0	=====	
TON/HR	0	0	12		
FT/MIN	120	120	120		
WIDTH, IN	12	12	12		
LENGTH, FT	20	20	20		
HP/CONVEYO	1.4	1.4	1.5		
HP	0.0	0.0	0.0		
\$/CONVEYOR	\$10,524	\$10,524	\$16,812		
\$ TOTAL	\$0	\$0	\$0		

APPENDIX E. CONSUMABLES AND WASTE RATES

TABLE E-1  
OPTIMIZED COOLSIDE PROCESS ON-SITE HYDRATION

Sorbent	Quick Lime (91 % CaO)		Annual Rates									
	Plant Capacity Factor	SO2 Removal	Removed Tons	SO2 Tons	Coal Sulfur %	Sorbent Tons	Soda Ash Tons	Water Mgal	Wastes (Dry) Tons	Fly Ash Credit Tons	Power MWh	
105	65%	70%	4367	7993	1.5	7993	1273	30525	29094	14508	8493	
158	65%	70%	6509	11915	1.5	11915	1898	45499	43368	21626	11053	
262	65%	70%	10575	19350	1.5	19350	3082	73919	70387	35075	15021	
530	65%	70%	21243	38869	1.5	38869	6192	148486	141390	70457	27658	
105	65%	70%	7275	15347	2.5	15347	2445	31143	41392	14195	8211	
158	65%	70%	10845	22877	2.5	22877	3644	46423	61699	21160	11012	
262	65%	70%	17619	37157	2.5	37157	5919	75417	100176	34328	15455	
530	65%	70%	35392	74639	2.5	74639	11889	151494	201229	68956	28432	
105	65%	70%	10182	24481	3.5	24481	3900	31704	56411	14102	8437	
158	65%	70%	15178	36492	3.5	36492	5813	47263	84088	21020	11301	
262	65%	70%	24658	59275	3.5	59275	9442	76776	136551	34105	15803	
530	65%	70%	49533	119070	3.5	119070	18967	154225	274298	68507	30534	

Hourly rate = annual rate/8760/capacity factor

# APPENDIX E. CONSUMABLES AND WASTE RATES

## TABLE E-2 OPTIMIZED COOLSIDE PROCESS WITH PURCHASED HYDRATE

Sorbent		Hydrated Lime (93 % CA(OH)2)		Annual Rates												
Plant Capacity Fac		65%		SO2 Removed		Soda Ash		Water		Wastes (Dry)		Fly Ash		Power		
SO2 Removal		70%		Tons		Tons		Mgal		Tons		Tons		MWh		
Plant Size	MW (gross)	Coal Sulfur %	SO2 Removed Tons	Sorbent Tons	Soda Ash Tons	Water Mgal	Wastes (Dry) Tons	Fly Ash Credit Tons	Power MWh							
105		1.5	4367	10330	1273	30524	29094	14508	7817							
158		1.5	6509	15398	1898	45500	43368	21626	10307							
262		1.5	10575	25006	3082	73919	70387	35075	14269							
530		1.5	21243	50231	6192	148486	141390	70457	26870							
105		2.5	7275	19834	2445	31143	41392	14195	7429							
158		2.5	10845	29565	3644	46423	61699	21160	10240							
262		2.5	17619	48019	5919	75417	100176	34328	14678							
530		2.5	35392	96459	11889	151494	201229	68956	27325							
105		3.5	10182	31638	3900	31704	56411	14102	7625							
158		3.5	15178	47161	5813	47263	84088	21020	10422							
262		3.5	24658	76604	9442	76776	136551	34105	14802							
530		3.5	49533	153879	18967	154225	274298	68508	29066							

Hourly rate = annual rate/8760/capacity factor

APPENDIX E. CONSUMABLES AND WASTE RATES

TABLE E-3  
OPTIMIZED LIMESTONE FORCED OXIDATION PROCESS

Sorbent Limestone (95 wt % CaCO <sub>3</sub> )		Annual Rates									
Plant Capacity Fac	65%	SO <sub>2</sub>		Soda Ash		Water	Wastes (Dry)		Fly Ash	Power	
SO <sub>2</sub> Removal	95%	Removed Tons	Sorbent Tons	Tons	Mgal	(Dry) Tons	Credit Tons	MWh			
Plant Size MW (gross)	Coal Sulfur %										
105	1.5	5927	10234	0	34146	16698	0	7893			
158	1.5	8834	15254	0	53877	24891	0	11524			
262	1.5	14352	24782	0	87528	40438	0	18383			
530	1.5	28829	49781	0	175842	81230	0	36286			
105	2.5	9874	17049	0	38913	27820	0	8946			
158	2.5	14718	25414	0	58010	41470	0	13095			
262	2.5	23911	41288	0	94224	67372	0	20911			
530	2.5	48032	82938	0	189269	135334	0	41321			
105	3.5	13819	23861	0	41646	38936	0	10117			
158	3.5	20599	35568	0	62076	58038	0	14797			
262	3.5	33465	57785	0	100852	94291	0	23637			
530	3.5	67223	116077	0	202593	189408	0	46931			

Hourly rate = annual rate/8760/capacity factor

**APPENDIX F**

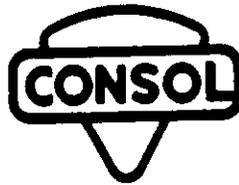
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**Pittsburgh, PA  
June 24-29, 1990**

Properties of Solid Wastes From the  
Edgewater Coolside and LIMB Process Demonstrations

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## INTRODUCTION

Waste management is an important element for the development of advanced SO<sub>2</sub> control processes. Consolidation Coal Company (Consol), a participant in the Coolside and LIMB process demonstrations at the Ohio Edison Edgewater Station (104 MWe), conducted a waste management study for the Coolside process in 1987-1988. Properties of Coolside waste were determined to address handling, transportation and landfill concerns for the solid waste which was to be generated during the Coolside process demonstration. The Coolside waste used in that study was from Consol's Coolside pilot plant (0.1 MWe). That study concluded that Coolside waste can be classified as a non-hazardous, solid waste, and can be disposed of by landfilling.<sup>1</sup> The Edgewater Coolside process demonstration, which began in July 1989 and ended in February 1990, was sponsored by the U.S. Department of Energy, the Ohio Coal Development Office, Ohio Edison, Babcock & Wilcox and Consol. Immediately prior to the Coolside tests, a demonstration test was conducted at the Edgewater station of the limestone injection multistage burner (LIMB) process. That program was funded by the U.S. Environmental Protection Agency. Waste samples were collected from both the LIMB and Coolside process demonstration tests for the present study.

In this paper, properties of solid wastes from the Edgewater Coolside and LIMB process demonstrations are discussed. Properties of Coolside waste generated in the demonstration program are compared with those reported previously for waste from the Coolside pilot plant.<sup>1</sup> The Coolside pilot plant waste data discussed here pertain to the material reported earlier<sup>1</sup> that was generated with NaOH additive. Both waste materials from the demonstration tests were evaluated for potential use in acid mine drainage (AMD) treatment and in the production of synthetic aggregate for road base and concrete applications. In the AMD treatment study, LIMB and Coolside wastes were used as the AMD neutralization reagents. Iron oxidation and neutralization rates, sludge settling properties and effluent water quality were compared with those obtained using conventional hydrated lime for treatment. The road base and concrete applications study included the preparation and characterization of synthetic aggregates from Coolside and LIMB wastes. The aggregates were prepared by pelletization and curing. Pelletization takes advantage of the pozzolanic properties of the wastes to produce a useful by-product. Pelletization also may benefit handling and transportation of the wastes and may reduce leachability if landfill disposal is required.

## EXPERIMENTAL

### SOURCE OF MATERIAL

The Coolside waste used in this study was taken from the ESP hopper of the Ohio Edison Edgewater station on October 6, 1989, during the Coolside process demonstration test.<sup>2</sup> The test was made under the following conditions: Ca/S mol ratio of 1.42, NaOH/Ca(OH)<sub>2</sub> weight ratio of 0.11, and 20°F approach to adiabatic saturation in the humidifier. Mississippi Lime Co. hydrated lime was the sorbent. Sulfur removal during the sampling period was about 56%. The LIMB waste was taken from the ESP hopper on May 25, 1989, during the LIMB process demonstration test. The test was made under the

following conditions: Ca/S mol ratio of 2.0, 150°F approach to adiabatic saturation in the humidifier. Marblehead hydrated lime was the sorbent. Sulfur removal during the sampling period was about 50%. Analyses of the Coolside and LIMB waste samples are listed in Table I. Both waste materials were collected during periods of stable operation at or near design conditions. As noted, the Coolside waste was obtained from operations at a Ca/S ratio of 1.42/1. Waste from operations at a Ca/S ratio of 2/1 would have been preferred for this study, but none was available. The lower Ca/S ratio (1.42/1 vs 2/1) caused the waste to contain more fly ash, less total Ca and, particularly, less unused  $\text{Ca}(\text{OH})_2$  than wastes produced at the higher Ca/S ratio. As a result, the comparison of properties of waste produced at Edgewater with waste from the Coolside pilot plant (Ca/S = 2/1) is influenced by these differences.

#### WASTE CHARACTERIZATION

The procedures used in this study to determine density, unconfined compressive strength, permeability coefficient and leachate characteristics were the same as reported previously.<sup>1</sup> The optimum moisture was measured based on the amount of water added to the dry-basis waste, as suggested in Reference 4. Particle size analysis was performed with a Malvern 2600C particle sizer. Particle size distribution was obtained in a dilute acetone slurry using the stirred cell attachment of the Malvern instrument.

#### AMD TREATMENT TESTS

Coolside and LIMB wastes were evaluated for use as substitutes for hydrated lime in AMD treatment. The wastes were subjected to laboratory test procedures originally established<sup>3</sup> to simulate the processing (preaeration, aeration and neutralization) in a Consol AMD treatment plant located in southwestern Pennsylvania. The raw AMD water, collected from the treatment plant just prior to testing, was stirred in a constant temperature bath at 13°C and preaerated with 4SCFH of air for 15 minutes. Following preaeration, a weighed quantity of Coolside or LIMB waste or Mercer hydrated lime was added as a 10% slurry to the AMD. Mercer hydrated lime is the commercial neutralization reagent used in the AMD treatment plant. The stirring and aeration were continued for 30 minutes after the addition of the reagent. pH and ferrous ion concentration were determined on aliquots (10 ml) before and after preaeration and at 10-minute intervals after the addition of the reagent. After 30 minutes aeration, the treated AMD was transferred to a one-liter graduated cylinder and allowed to settle. The sludge settling rate was observed for one hour. After 24 hours, the clarified supernatant was sampled for analyses. After 24 hours of settling, the composition of the supernatant water approximates that of the water that would be discharged from the AMD treatment plant. The ferrous ion concentration was measured during the test by potassium dichromate titration after filtration. Analyses of trace and major elements in the supernatant were performed as reported previously.<sup>1</sup>

#### WASTE PELLETIZATION TEST

Coolside and LIMB wastes were pelletized by a two-step method. The waste was first mixed with the appropriate amount of water in an Eirich

mixer. The wetted material was then pelletized with an additional amount of water in a rotary disc pelletizer (16-inch ID). The pellets were cured at ambient temperature and humidity in the laboratory, and at 100°F and 100% humidity in a curing box, separately, for strength testing. Strength tests, including both drop strength and compressive strength, were performed at various curing times. The drop strength test was conducted by dropping ten ca. 1/4 x 1/2-inch pellets six feet onto a concrete floor and measuring the weight percent remaining larger than 1/4 inch. The compressive strengths of individual pellets were measured with a Soiltest compressive strength apparatus. The compressive strength values reported in this paper are each the average of ten measurements on ca. 1/4 x 1/2-inch pellets. Other properties of the pellets were determined according to Standard ASTM methods. The LA Abrasion Index and Soundness Index of the LIMB waste pellets were performed by West Penn Testing Laboratory, Pittsburgh, Pennsylvania.

## RESULTS AND DISCUSSION

### WASTE CHARACTERIZATION

The compositions of Coolside and LIMB wastes (Table I) show differences reflecting differences in the feed sorbent and differences in process chemistry. The unused sorbent is in the form of hydrated lime ( $\text{Ca}(\text{OH})_2$ ) in Coolside waste and in the form of quicklime ( $\text{CaO}$ ) in LIMB waste. The major component of the spent sorbent is calcium sulfite ( $\text{CaSO}_3$ ) in Coolside waste and calcium sulfate ( $\text{CaSO}_4$ ) in LIMB waste. Coolside waste also contains small amounts of  $\text{Na}_2\text{SO}_3$  and  $\text{Na}_2\text{SO}_4$  as minor components. The balance of both materials is fly ash. These compositional differences can make the physical and chemical properties of these two wastes quite different. For instance, LIMB waste exhibited a substantial temperature rise (ca. 150°F) upon wetting, caused by the exothermic reaction of water and  $\text{CaO}$  (unused sorbent) to produce  $\text{Ca}(\text{OH})_2$ . In contrast, Coolside waste, in which the unused sorbent is  $\text{Ca}(\text{OH})_2$ , exhibited only a slight increase in temperature (10°F) upon wetting.<sup>1</sup> The amount of temperature rise upon wetting can affect handling and transportation of the waste.

In this study, the following properties of Edgewater Coolside and LIMB wastes were determined: density and particle size, strength and permeability, and leachate characteristics. Density and particle size distribution are relevant to handling, transportation and disposal. Strength, permeability and leachate characteristics are most relevant to landfill disposal.

### Density and Particle Size

The following table lists various properties of Edgewater Coolside and LIMB wastes.

	<u>Coolside Waste</u>	<u>LIMB Waste</u>
<u>Waste, as received</u>		
Bulk density (g/mL)		
Loose	0.47	0.53
Tapped	0.62	0.76

	<u>Coolside Waste</u>	<u>LIMB Waste</u>
<u>Waste, as received</u>		
Particle size ( $\mu\text{m}$ )		
Mean diameter	13.6	9.9
Diameter below which 90% sample lies	34.8	20.2
Diameter below which 50% sample lies	7.6	6.8
Diameter below which 10% sample lies	3.0	3.0
Moisture content		
(wt %, as determined)	0.64	0.06
<u>Waste, as compacted (ASTM D698)</u>		
Optimum moisture content		
(wt % added water, dry basis)	33	46
Maximum dry bulk density (g/mL / lb/ft <sup>3</sup> )	1.31/81.8	1.33/83.0
Moisture content, measured (wt %, dry basis)	31.2	31.8

The Coolside waste has lower bulk densities (loose and tapped) and a wider particle size distribution than the LIMB waste. The higher optimum moisture value of the LIMB waste was caused by consumption of water in the hydration reaction and by water evaporation caused by the temperature rise upon wetting. The optimum moisture content is reported as the wt % of added water.<sup>4</sup> Moisture contents measured after compaction were essentially the same for both materials. The maximum dry bulk density, determined at the optimum moisture content, was about the same for the two wastes. These data are important for landfill disposal of these wastes.

There were significant differences in maximum dry bulk density of Coolside waste from the Edgewater demonstration test and from the Coolside pilot plant (81.8 lb/ft<sup>3</sup> vs 66.8 lb/ft<sup>3</sup>). Chemical compositions and particle size distributions, which can both affect the maximum dry bulk density, are listed below for both materials.

	<u>Edgewater Coolside Waste</u>	<u>Pilot Plant Coolside Waste</u>
Maximum dry bulk density (lb/ft <sup>3</sup> )	81.8	66.8
<u>Chemical compositions</u>		
<u>wt % of waste, dry basis</u>		
Sulfite sulfur	4.9	5.6
Sulfate sulfur	1.2	2.2
Ca(OH) <sub>2</sub> , wt	22	26
<u>Particle size (<math>\mu\text{m}</math>)</u>		
Mean diameter	13.6	7.2
Diameter below which 90% sample lies	34.8	16.7
Diameter below which 50% sample lies	7.6	5.5
Diameter below which 10% sample lies	3.0	2.3

The Edgewater and pilot plant Coolside wastes were generated at Ca/S mol ratios of 1.42 and 2.0, respectively. The Edgewater Coolside waste contains less spent and unused sorbents (i.e., less CaSO<sub>3</sub>/CaSO<sub>4</sub> and Ca(OH)<sub>2</sub> and more

fly ash) because of the lower Ca/S ratio and it has a wider particle size distribution than the pilot plant Coolside waste. Both of these factors give the Edgewater waste a higher maximum dry bulk density.

### Strength and Permeability

Unconfined compressive strength is a good indication of structural integrity during landfill disposal for dry, lime-enriched FGD wastes that have pozzolanic properties.<sup>5</sup> Permeability measurements, along with extraction data, can be used to evaluate possible effects of leachate on groundwater quality. The unconfined compressive strength of a landfilled waste should be high enough to maintain its structural integrity. Landfilled materials may be subjected to numerous passes of earth-moving equipment which may exert pressures up to 19 psi. Permeability should be low in order to minimize the passage of leachate through the disposed waste in a landfill. Permeability is typically reported as a permeability coefficient (cm/sec). FGD wastes with permeability coefficients of  $10^{-6}$  cm/sec and lower are generally considered to be suitable for landfill disposal.<sup>6</sup> The following table summarizes the unconfined compressive strengths and permeability coefficients of the wastes.

	<u>Coolside Waste</u>		<u>LIMB Waste</u>
Added water (wt % dry basis)	33	31	46
Dry bulk density (lb/ft <sup>3</sup> )	81.8	77.5	83.0
<u>Unconfined compressive strength (psi)</u>			
0 days curing	19	41	-
7 days curing	135	106	73
14 days curing	422	276	144
28 days curing	655	524	489
<u>Permeability coefficient (cm/sec)</u>			
0 days curing	$1.2 \times 10^{-6}$	-	$8.8 \times 10^{-6}$
7 days curing	$1.7 \times 10^{-7}$	-	$7.0 \times 10^{-7}$
14 days curing	$5.6 \times 10^{-8}$	-	$2.0 \times 10^{-7}$
28 days curing	$1.7 \times 10^{-8}$	-	$2.0 \times 10^{-7}$

Unconfined compressive strength and permeability coefficient were measured according to ASTM Method D-1633 and Earth Method 1110-02-1906, respectively, on waste samples prepared according to ASTM D-698 and cured according to ASTM Method C-192.

As expected, the unconfined compressive strength increased and the permeability coefficient decreased with increasing curing time, illustrating the pozzolanic properties of both LIMB and Coolside wastes. For Coolside waste, unconfined compressive strength also was determined with reduced water addition. The dry bulk density of that sample was equivalent to 95% of the maximum dry bulk density, which is obtained at the optimum moisture. The less dense material gave slower strength development, showing the effect of density on strength development. These data are reported because landfilled wastes usually have densities a few percent less than the maximum dry bulk density.<sup>6</sup>

The Edgewater Coolside waste has a higher unconfined compressive strength and a lower permeability coefficient than the pilot plant waste as shown below (tests were conducted at optimum moisture content). These differences are associated with the higher maximum dry bulk density of the Edgewater Coolside waste which, in turn, results from the lower Ca/S ratio used during its production. Based on the data shown below, unconfined compressive strength was suitably high and the permeability was low enough in both Edgewater Coolside and LIMB wastes to permit landfill disposal, as also indicated in the previous study with pilot plant waste.<sup>1</sup>

	<u>Edgewater Coolside Waste</u>	<u>Pilot Plant Coolside Waste</u>
Maximum dry bulk density (lb/ft <sup>3</sup> )	81.8	66.8
Optimum moisture content (wt %, dry basis)	33	30
<u>Unconfined compressive strength (psi)</u>		
0 days curing	19	32
7 days curing	135	88
14 days curing	422	159
28 days curing	655	251
<u>Permeability coefficient (cm/sec)</u>		
0 days curing	$1.2 \times 10^{-6}$	$1.2 \times 10^{-5}$
7 days curing	$1.7 \times 10^{-7}$	$1.4 \times 10^{-6}$
14 days curing	$5.6 \times 10^{-8}$	---
28 days curing	$1.7 \times 10^{-8}$	$2.0 \times 10^{-7}$

#### Leachate Characteristics

Leachate toxicity tests were performed on both Edgewater Coolside and LIMB wastes, according to the EP toxicity test procedures. The leachate concentrations listed in Table II include the trace elements of the primary and secondary drinking water standards, major anions and cations, and total dissolved solids (TDS). Trace elements were analyzed because the concentrations of these elements may be subjected to state and local governmental regulations. Good material balances (based on measured and calculated TDS) and charge balances (based on concentrations of anions and cations) were obtained in both leachates, when the acetate concentration in the leachate was taken into consideration as in the previous study.<sup>1</sup> As shown in Table II, the sodium concentration was higher and the calcium concentration was lower in the leachate from Coolside waste than from the LIMB waste. The sodium and calcium concentrations were 1222 and 2583 mg/L, respectively, in the Coolside waste leachate and 20.2 and 3417 mg/L, respectively, in the LIMB waste leachate. Generally, leachate sodium and calcium concentrations are not regulated. However, TDS, which includes sodium and calcium concentrations, may be subject to state and local governmental regulations. As indicated in Table II, the TDS value was only slightly higher in the Coolside waste leachate (11920 mg/L) than in the LIMB waste leachate (11814 mg/L). Both were slightly lower than the TDS value (12598 mg/L) in the leachate from pilot plant Coolside waste.<sup>1</sup> As reported for the pilot

plant Coolside waste, the RCRA elements (As, Ba, Ca, Cr, Pb, Hg, Ag and Se) in the leachates from the Edgewater wastes are well below allowable limits. Both Edgewater Coolside and LIMB wastes can be classified as non-hazardous.

The initial and final pHs of Edgewater Coolside and LIMB wastes in the EP toxicity test are listed below. Initial pH was measured before the addition of acetic acid.

	pH	
	<u>Initial</u>	<u>After 24 Hours</u>
Coolside waste	12.45	12.20
LIMB waste	12.43	12.20

As with other lime-enriched FGD wastes,<sup>7</sup> run-off water may need to be neutralized before discharge, depending on the site-specific situation of a landfill.

#### AMD Treatment

Both LIMB and Coolside wastes were evaluated as neutralization reagents for AMD treatment in laboratory tests. Iron oxidation and neutralization rates, sludge settling properties and effluent water quality were examined. Results were compared with those obtained using conventional hydrated lime treatment. These tests were conducted with the raw AMD water, collected just prior to the testing, from a Consol AMD treatment plant in southwestern Pennsylvania. Analyses of the raw and treated AMD water are shown in Tables III and IV for LIMB waste and Coolside waste, respectively. The hydrated lime used was the commercial neutralization reagent (Mercer hydrated lime) used in the AMD treatment plant.

Figures 1 and 2 show the changes in the fraction of ferrous ion remaining and pH, respectively, of the treated AMD during preaeration and neutralization with LIMB waste. At a dosage of 1.40 g/L to 1.60 g/L of AMD water, LIMB waste produced results similar to hydrated lime at a dosage of 0.40 g/L of AMD water in terms of its ability to promote iron oxidation and neutralize the AMD water. The high dosage of LIMB waste was required because of its low lime index relative to the Mercer hydrated lime (23% vs 91%, as  $\text{Ca}(\text{OH})_2$ ). Figure 3 illustrates the sludge settling rates during the LIMB waste and hydrated lime tests. These settling tests were performed in one-liter graduated cylinders immediately after neutralization. The sludge produced from treating the AMD water with LIMB waste settled at a slightly faster rate than the sludge produced from treating the AMD water with hydrated lime, even though the dosage was 3.5 to 4 times higher with the LIMB waste. This was probably due to the fly ash component of the LIMB waste which may act as precipitation seeds on which the sludge may accumulate, thus enhancing the settling behavior. Magnetite (a component in fly ash) is known to be able to enhance the settling characteristics of AMD sludge.<sup>8</sup> The sludge volume after 24 hours was lower with the LIMB waste, as indicated in Figure 3. The high dosage of LIMB waste used in the AMD treatment did not cause problems such as slow settling and high sludge volume based on these laboratory tests. Table III shows the analyses of the supernatant produced by LIMB waste and hydrated lime treatments after the sludge had settled for

24 hours. NPDES permit limits also are shown for comparison. All parameters are within the NPDES permit limits for the LIMB waste treatment. Other parameters with LIMB waste treatment also are similar to the results obtained with hydrated lime treatment, indicating comparable discharge water quality in both treatments. In summary, LIMB waste appears to be a suitable substitute for hydrated lime as an AMD treatment reagent, even though considerably higher reagent dosages are required.

Figures 4 and 5 show the changes in the fraction of ferrous ion remaining and pH, respectively, in AMD treated with Coolside waste and hydrated lime. Coolside waste at dosages of 1.05 g/L and 1.21 g/L produced results similar to those with 0.31 g/L hydrated lime. The raw AMD used in these tests (Table IV) was of higher quality than that used for the LIMB waste tests (Table III), and thus reagent dosages are lower. Coolside waste has a lime index of 22% (as  $\text{Ca(OH)}_2$ ), similar to that of the LIMB waste. As in the LIMB waste tests, Coolside waste produced comparable results to hydrated lime with regard to iron oxidation and neutralization (Figure 4 and 5). The Coolside waste treatment produced a more compacted sludge than hydrated lime treatment and the sludge settled more quickly (Figure 6). As with LIMB waste treatment, Coolside waste treatment did not result in slow settling or high sludge volume in the laboratory test. As indicated in Table IV, water quality also was maintained and meets all NPDES guidelines. One issue that still must be addressed for the use of Coolside waste as an AMD reagent is the possibility that, over time, the sulfite in the waste will reduce the precipitated ferric hydroxide in the AMD sludge and thus remobilize it.

### Waste Pelletization

In this study, strength, density, particle size and leachability of pelletized Edgewater LIMB and Coolside wastes were evaluated. These properties are relevant to handling, transportation, disposal and utilization.

The strength development of waste pellets was dependent on curing conditions. For instance, the drop strength of Coolside waste pellets was enhanced greatly with increasing curing temperature from ambient to 100°F, as shown in Figure 7. A similar effect was also observed with LIMB waste pellets. Drop strength is important for handling and transportation.<sup>9</sup> These data indicated that pelletization and adequate curing can enhance waste handleability for transportation.

Table V lists properties of LIMB and Coolside waste pellets cured at 100°F with 100% humidity for 28 days. The cured waste pellets were evaluated for potential use in the production of synthetic aggregate for road base and concrete applications. For LIMB waste pellets, the LA abrasion index and density (dried) were within the requirements for lightweight aggregate; however, the soundness index was too high for this use. Freeze-thaw tests will be performed to confirm the soundness test results. The particle size of the pellets complied with the size specification of the most popular aggregate products, i.e., AASATO 57 and 67. It appears that LIMB waste pellets are suitable for use as synthetic aggregate in road base applications (Class C). Insufficient data are available for evaluation of Coolside waste pellets as synthetic aggregate.

Leachate toxicity tests were performed on LIMB and Coolside waste pellets (Table VI). Both pelletized products can be classified as non-hazardous. It is noteworthy that the sodium concentration in the Coolside leachate decreased markedly from 1222 mg/L (Table II) to 399 mg/L (Table VI) after pelletization and curing. This indicates that pelletization is a feasible way to reduce sodium leachability of Coolside waste.

#### CONCLUSIONS

Properties of Coolside and LIMB wastes from the Edgewater (104 MWe) demonstration program were characterized. The properties determined included density and particle size, strength and permeability, and leachate characteristics. The tests results indicated that both wastes are suitable for landfill disposal. The Edgewater Coolside waste had a higher compressive strength and a lower permeability coefficient than that of the Coolside waste produced in the pilot program. These differences were associated with the higher maximum dry bulk density of the Edgewater Coolside waste which, in turn, resulted in part from the lower Ca/S ratio in use during sampling at the Edgewater plant. Coolside and LIMB wastes were also evaluated for use in acid mine drainage (AMD) treatment. Both wastes appear to be suitable replacements for conventional hydrated lime in AMD treatment in terms of iron oxidation and neutralization rates, sludge settling properties and effluent water quality. LIMB and Coolside wastes were pelletized and the products were characterized in terms of strength, density, particle size and leachability. Pelletization enhances handleability for transportation and reduces waste leachability. The pelletized products, after curing, have potential to be used as synthetic aggregates for road base construction.

#### ACKNOWLEDGEMENT

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Table I. Analyses of LIMB and Coolside wastes from the Edgewater demonstration project.

	<u>LIMB Waste</u>	<u>Coolside Waste</u>
Moisture, wt % of waste (as received)	0.06	0.64
<u>Ultimate Analysis, wt % of Waste (dry basis)</u>		
Carbon	2.84	4.44
Hydrogen	0.09	0.93
Nitrogen	0.09	0.09
Sulfur (total)	4.80	6.05
Sulfate sulfur	4.42	1.16
Sulfite sulfur	0.02	4.88
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	3.07	1.73
Ash	95.94	89.85
<u>Elemental Analysis, wt % of Waste (dry basis)</u>		
SiO <sub>2</sub>	26.02	20.94
Al <sub>2</sub> O <sub>3</sub>	11.11	9.79
TiO <sub>2</sub>	0.50	0.44
Fe <sub>2</sub> O <sub>3</sub>	12.72	8.88
CaO	30.65	28.52
MgO	0.74	0.61
Na <sub>2</sub> O	0.12	2.10
K <sub>2</sub> O	1.05	0.95
P <sub>2</sub> O <sub>5</sub>	0.21	0.10
SO <sub>3</sub>	12.29	15.80
Unaccounted	4.68	11.87
Lime Index (as wt % Ca(OH) <sub>2</sub> )	23	22

Table II. EP test leachate compositions of Edgewater  
Coolside and LIMB wastes.

<u>Analysis</u>	<u>Concentration in Leachate, mg/L</u>		<u>RCRA Limit</u>
	<u>Coolside Waste</u>	<u>LIMB Waste</u>	
Hydroxide as CaCO <sub>3</sub>	1520	1810	-
Carbonate as CaCO <sub>3</sub>	560	0	-
Total dissolved solids	11920	11814	-
Calcium	2583	3417	-
Sodium	1222	20.2	-
Arsenic	0.0331	0.0490	5
Barium	0.973	1.397	100
Cadmium	0.00216	0.00131	1
Chromium	0.0717	0.0031	5
Lead	0.0052	0.00966	5
Mercury	<0.020	<0.020	0.2
Silver	<0.010	<0.010	5
Selenium	0.117	0.0525	1
Nitrate (as N)	<0.001	<0.001	-
Chloride	275	100	-
Sulfate	1354	1579	-
Iron	<0.20	<0.20	-
Zinc	0.0204	0.0199	-
Manganese	0.20	0.20	-
Copper	0.0113	0.0185	-

Table III. Comparison of AMD water quality before and after LIMB waste  
and hydrated lime treatments with NPDES limits.

<u>Parameter</u>	<u>Raw AMD Water (a)</u>	<u>One Liter Laboratory Test</u>			<u>NPDES Permit Limits</u>
		<u>1.4 g/L LIMB Waste</u>	<u>1.6 g/L LIMB Waste</u>	<u>0.4 g/L Hydrated Lime</u>	
pH	6.43	8.28	8.32	8.27	6-9
Acidity (CaCO <sub>3</sub> )	214	-123	-143	-144	-
Alkalinity (CaCO <sub>3</sub> )	156	144	153	158	-
Aluminum	<0.05	<0.05	<0.05	<0.05	-
Calcium	258	464	467	417	-
Iron, total	320	1.53	1.55	0.22	2.5
Magnesium	99.2	100	98.5	97.3	-
Manganese	4.97	1.06	0.69	0.34	2.0
Potassium	8.11	8.69	9.23	13.6	-
Sodium	1671	1635	1611	1586	-
Osmotic pressure	102	113	111	109	170 (b)
Chromium	<0.05	<0.05	<0.05	<0.05	-
Chloride	535	520	525	530	-
Sulfate	3702	3999	3978	3793	-

a) Collected on June 19, 1989.

b) Applies to this plant only.

Table IV. Comparison of AMD water quality before and after Coolside waste and hydrated lime treatment with NPDES limits.

Parameter	Raw AMD Water (a)	One Liter Laboratory Test			NPDES Permit Limits
		1.05 g/L LIMB Waste	1.21 g/L LIMB Waste	0.31 g/L Hydrated Waste	
pH	6.78	8.69	8.98	8.50	6-9
Acidity (CaCO <sub>3</sub> )	102	-159	-103	-249	-
Alkalinity (CaCO <sub>3</sub> )	262	170	117	260	-
Aluminum	<0.22	<0.200	<0.200	<0.200	-
Calcium	204	248	273	264	-
Iron, total	232	0.56	0.62	1.04	2.5
Magnesium	77.8	75.6	77.7	77.6	-
Manganese	2.97	<0.20	<0.20	<0.20	2.0
Potassium	7.88	11.2	9.16	10.1	-
Sodium	1562	1557	1590	1578	-
Osmotic pressure	117	106	108	109	170 (b)
Chromium	<0.001	<0.001	<0.001	<0.001	-
Chloride	570	545	540	534	-
Sulfate	3130	3305	3353	3202	-

a) Collected on January 24, 1990.

b) Applies to this plant only.

Table V. Properties of LIMB and Coolside waste pellets.

	LIMB Waste Pellets	Coolside Waste Pellets
<u>Product Strength</u>		
LA abrasion index, wt % (ASTM C-131)	37.8	-
Soundness index, wt % (ASTM C-88)	47.0	-
Pellet compressive strength, lbs	80.9	41.2
Pellet drop strength, wt %	100	100
<u>Product Density</u>		
Unit weight (ASTM C-29), lb/ft <sup>3</sup>		
Wet	67.2	61.4
Dry	46.0	43.7
<u>Product Size Distribution, wt %</u>		
+ 1"	1.31	0.55
1 x 3/4"	8.26	7.90
3/4 x 1/2"	32.89	43.24
1/2 x 3/8"	30.96	19.16
3/8 x 1/4"	12.28	8.38
1/4" x 4 mesh	6.53	5.00
4 x 8 mesh	6.41	10.38
- 8 mesh	1.36	5.39

Table VI. EP leachate compositions of Edgewater Coolside and LIMB waste pellets.

Analysis	Concentration in Leachate, mg/L	
	Coolside Waste Pellets	LIMB Waste Pellets
Hydroxide as CaCO <sub>3</sub>	0	70
Carbonate as CaCO <sub>3</sub>	70	0
Total dissolved solids	10902	9648
Calcium	2289	2415
Sodium	399	21
Arsenic	0.0145	0.0285
Barium	0.123	0.230
Cadmium	0.00185	0.00187
Chromium	0.00716	0.046
Lead	<0.001	<0.001
Mercury	<0.020	<0.020
Silver	<0.010	<0.010
Selenium	0.0909	0.0145
Nitrate (as N)	<0.001	<0.001
Chloride	74	68
Sulfate	1668	1164
Iron	<0.05	<0.05
Zinc	0.0183	0.0140
Manganese	<0.05	<0.05
Copper	0.00522	0.00300

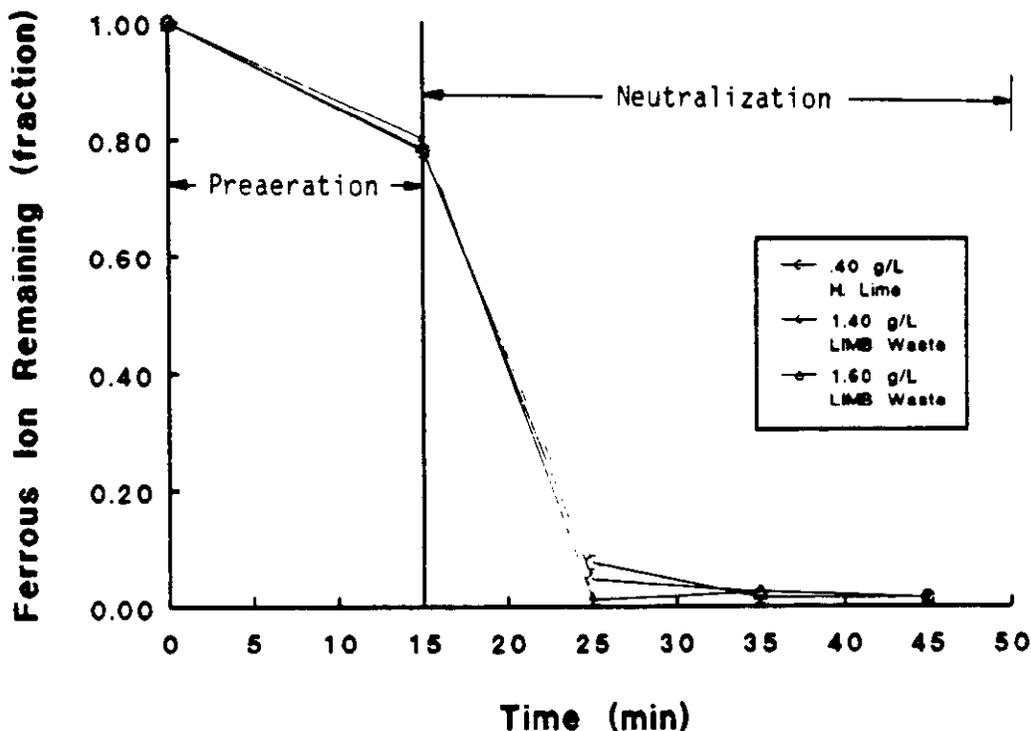


Figure 1. Rates of iron oxidation in AMD treated with LIMB waste and hydrated lime.

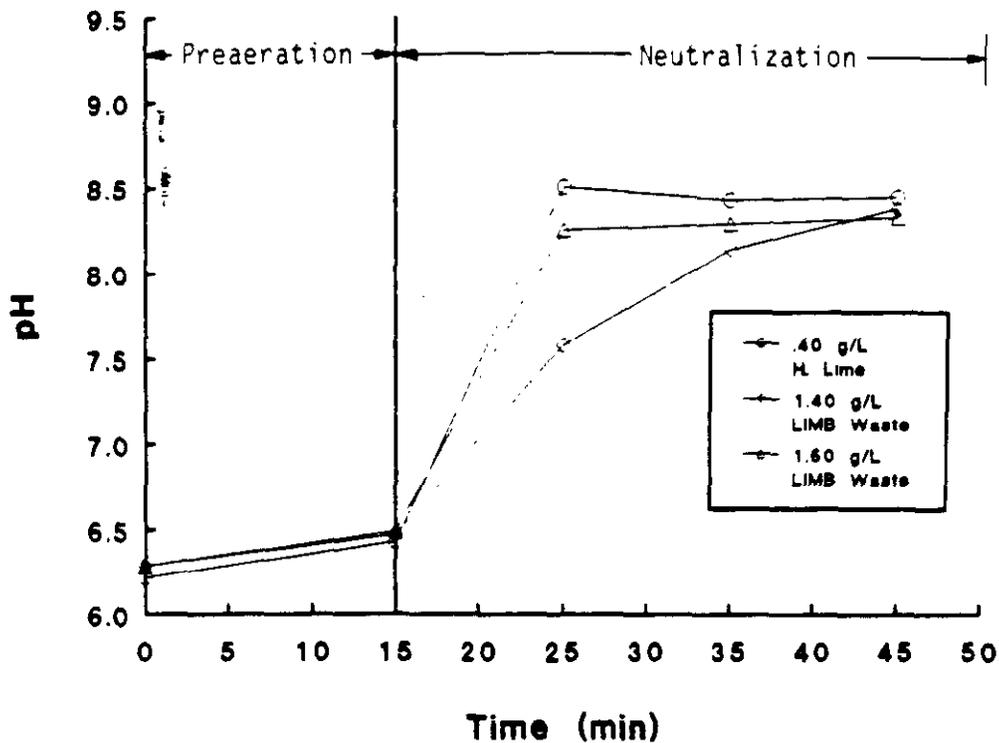


Figure 2. Changes in pH with time for AMD treated with LIMB waste and hydrated lime.

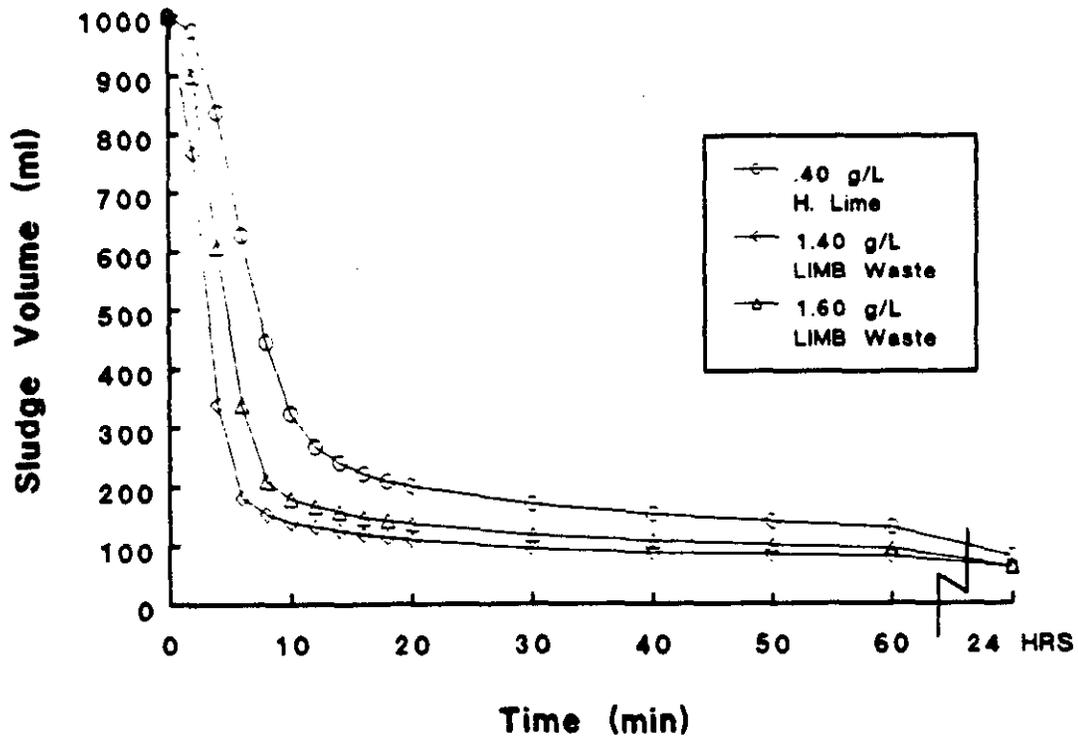


Figure 3. Sludge settling rates for AMD treated with LIMB waste and hydrated lime.

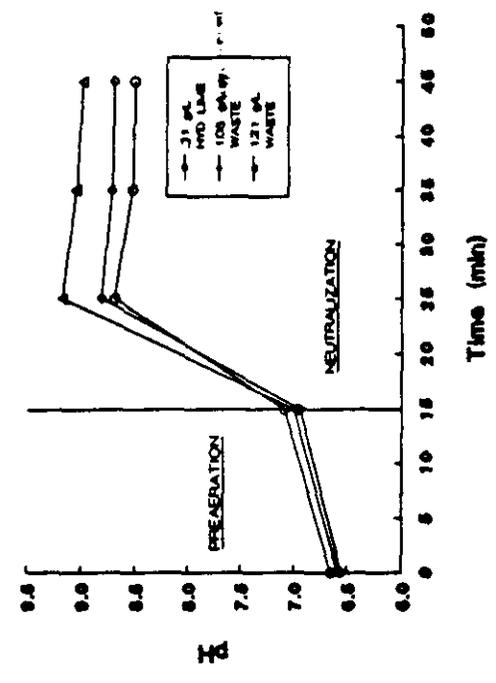


Figure 4. Rates of iron oxidation in AMD treated with Coolside waste and hydrated lime.

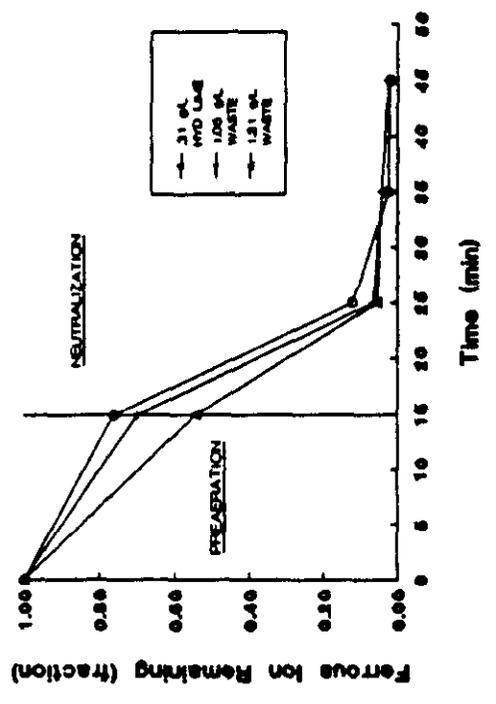


Figure 5. Changes in pH with time for AMD treated with Coolside waste and hydrated lime.

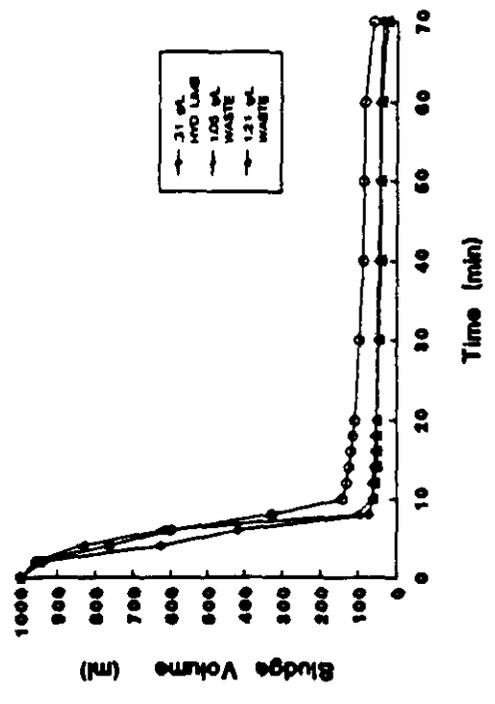


Figure 6. Sludge settling rates for AMD treated with Coolside waste and hydrated lime.

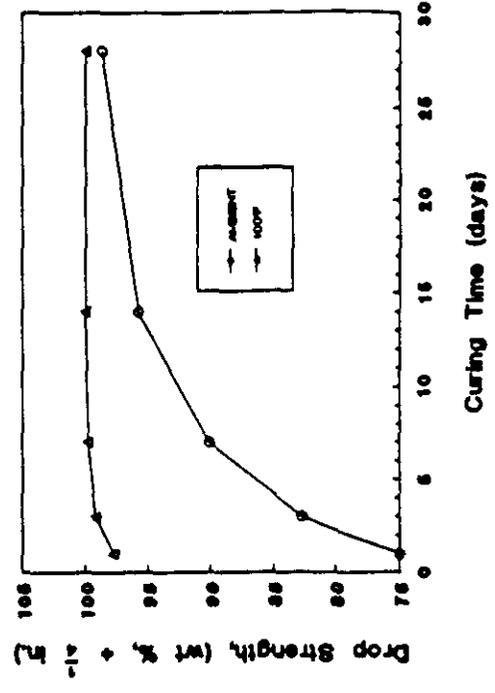


Figure 7. Drop strength of Coolside waste pellets vs. curing temperature and time.

## APPENDIX G. INSTRUMENTATION NOMENCLATURE

AIC	Analyzer Indicating Controller
FE	Flow Element
FIC	Flow Indicating Controller
FR	Flow Recorder
FRC	Flow Recording Controller
LIC	Level Indicating Controller
PI	Pressure Indicator
PR	Pressure Recorder
PRC	Pressure Recording Controller
PT	Pressure Transmitter
RO	Restriction Orifice
SC	Speed Controller
SP	Set point
TC	Thermocouple
WT	Weight Transmitter
XCV	Unclassified Control Valve

## APPENDIX H: METRIC CONVERSION TABLE

<u>To convert from</u>	<u>To</u>	<u>Multiply by</u>
in	m	$2.540 \times 10^{-2}$
ft	m	$3.048 \times 10^{-1}$
lb	kg	$4.536 \times 10^{-1}$
lb/10 <sup>6</sup> Btu	ng/J	$4.299 \times 10^2$
cfm	m <sup>3</sup> /s	$4.719 \times 10^{-4}$
gr/cf	kg/m <sup>3</sup>	$2.288 \times 10^{-3}$
ft <sup>2</sup> /1000 cfm	m <sup>2</sup> /1000 m <sup>3</sup> /s	$1.968 \times 10^{-2}$
micron	m	$1.000 \times 10^{-6}$
ton	kg	$9.072 \times 10^2$
mile	km	1.609
Btu	J	$1.055 \times 10^3$
ft <sup>3</sup>	m <sup>3</sup>	$2.832 \times 10^{-2}$
gal	m <sup>3</sup>	$3.785 \times 10^{-3}$
acre	m <sup>2</sup>	$4.047 \times 10^3$
ft/s	m/s	$3.048 \times 10^{-1}$
lb/ft <sup>3</sup>	kg/m <sup>3</sup>	$1.602 \times 10^1$
mcfm	m <sup>3</sup> /sec	$4.719 \times 10^{-1}$
μA/ft <sup>2</sup>	μA/m <sup>2</sup>	$1.076 \times 10^1$
°F	°C	°C = (5/9)(°F-32)
lb/h	kg/h	$4.536 \times 10^{-1}$
psi	kPa	6.895
Btu/lb	kJ/kg	2.326