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To
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From
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FINAL REPORT - OCDO ADVANCED DRY SCRUBBING PROJECT	DECEMBER 23, 1994

This letter to cover one customer and one subject only

Attached is the Final Report for Project 43117, Development of Advanced, Dry, SO_x/NO_x Emission Control Technologies for High-Sulfur Coal. Please transmit the required number of copies to OCDO with the following items included in an appropriate cover letter as required by the OCDO Final Report Format.

A. **Project Title:** Development of Advanced, Dry, SO_x/NO_x Emission Control Technologies for High-Sulfur Coal

OCDO Grant Number: CDO/D-922-15

B. **Grantee:** Babcock & Wilcox
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Advanced Dry Scrubbing
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A+LIDS
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Furnace Ammonia Injection
Droplet Impingement Device
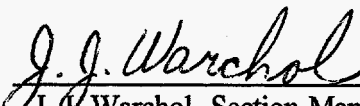Economic Analysis

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D.   **Final Report**

E.   **Sponsors:** This project was funded in part by the Ohio Coal Development Office, Department of Development, State of Ohio, and the Babcock & Wilcox Co.

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G. T. Amrhein

Reviewed and Approved by: J. J. Warchol, Section Manager

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Development of Advanced, Dry, SO_x/NO_x
Emission Control Technologies
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1.0 <u>EXECUTIVE SUMMARY</u>

Dry Scrubbing is a common commercial process that has been limited to low- and medium-sulfur coal applications because high-sulfur coal requires more reagent than can be efficiently injected into the process. Babcock & Wilcox has made several advances that extend dry scrubbing technologies to higher sulfur coals by allowing deposit-free operation at low scrubber exit temperatures. This not only increases the amount of reagent that can be injected into the scrubber, but also increases SO_2 removal efficiency and sorbent utilization.

The objectives of this project were to demonstrate, at pilot scale, that advanced, dry-scrubbing-based technologies can attain the performance levels specified by the 1990 Clean Air Act Amendments for SO_2 and NO_x emissions while burning high-sulfur coal, and that these technologies are economically competitive with wet scrubber systems. The use of these technologies by utilities in and around Ohio, on new or retrofit applications, will ensure the future of markets for high-sulfur coal by creating cost effective options to coal switching.

1.1   <u>Project Results and Conclusions</u>

All tests were conducted in B&W's pilot-scale furnace known as the Small Boiler Simulator (SBS). The SBS is PC-fired and rated at 5 million Btu/hr. Other test equipment included a 5 ft diameter, vertical, down-flow dry scrubber with dual-fluid atomization, and a pulse-jet baghouse. The project evaluated four advanced dry-scrubbing-based technologies in various stages of development:

Advanced Dry Scrubbing
LIDS - Limestone Injection Dry Scrubbing
A+LIDS - Ammonia and Limestone Injection Dry Scrubbing
Magnesium-Based Dry Scrubbing

The final test phase was completed in August. A separate Milestone Report was issued for each technology and the results are presented below.

Advanced Dry Scrubbing

In the B&W dry scrubbing process, pebble lime (CaO) is slaked to produce a highly reactive slurry. The slurry is atomized with high-pressure air and injected into a dry scrubber to remove SO_2. The scrubber is designed so that the slurry droplets dry before reaching the walls or exit. Unreacted sorbent, flyash, and reaction products are collected in a baghouse. Additional SO_2 removal occurs across the baghouse filter cake because of the extended contact time with the flue gas and efficient cross-flow contact. A portion of the baghouse waste material can be mixed with the fresh sorbent to increase sorbent utilization or to increase the solids concentration of the slurry to improve drying characteristics.

The objectives of this phase were to establish a database for comparison to wet scrubbing and other dry scrubbing technologies, and to demonstrate B&W's advances in dry scrubbing. The performance goal was 90-95% SO_2 removal at a Ca/S ratio less than 1.5 using a high-sulfur Ohio Coal. To achieve these goals, B&W's advancements were used to operate the dry scrubber and baghouse at approach to saturation temperature (T_{as}) of 10°F and less.

Tests were carried-out using a medium-sulfur, washed, Ohio #6 coal containing 1.32% sulfur that produced about 1000 ppm of SO_2 at the scrubber inlet, and a high-sulfur Ohio #6 containing 4.35% sulfur that produced about 3600 ppm of SO_2 at the scrubber inlet.

Performance goals were exceeded by operating the scrubber at low T_{as}. At a Ca/S ratio of 1.15 and a T_{as} of 10°F, over 99% SO_2 removal was attained. Calcium utilization for several tests was above 90%. Coal sulfur had no effect on performance, but the use of high-sulfur coal limited the Ca/S ratio to a maximum of 1.2.

Limestone Injection Dry Scrubbing (LIDS)

LIDS combines furnace limestone injection and dry scrubbing to achieve high levels of SO_2 removal and reduce reagent costs while burning high-sulfur coal. In the process, pulverized limestone is injected into the upper furnace and reacts with a portion of the SO_2. Unreacted lime and waste material pass through the system and are collected in a baghouse. A portion of this material is slaked to produce a highly reactive slurry for dry scrubbing. The slurry is atomized with high-pressure air and injected into the dry scrubber to remove more SO_2. Additional SO_2 removal occurs in the baghouse. The differences between LIDS and dry scrubbing are that SO_2 removal occurs in **three** stages (furnace, scrubber, baghouse), and that the waste ash from limestone furnace injection is the only source of lime for the dry scrubber, no pebble lime is used.

The objective of this phase was to build on the strong data base attained during the OCDO sponsored Pilot-Scale LIDS Demonstration in 1990-91[1,2]. Results from 1991 showed that 92% SO_2 removal was possible at a Ca/S ratio of 1.9. The goal of this phase was 95% SO_2 removal at a Ca/S ratio less than 1.5.

Tests were conducted with an Ohio #6 coal containing 3.94% sulfur that produced about 3000 ppm of SO_2 at the scrubber inlet. Conditions included furnace Ca/S ratios ranging from 1.00 to 2.09, nominal T_{as} from 10°F to 20°F, and the addition of 1% $CaCl_2$ to the slurry solids to increase the moisture content of the baghouse ash to about 1.0%. Performance goals were exceeded in tests that achieved over 99% SO_2 removal at a Ca/S of 1.44 and a T_{as} of 11°F.

Ammonia and Limestone Injection Dry Scrubbing (A+LIDS)

A+LIDS, (US Patent 5,176,088) is a new dry-scrubber-based technology developed by B&W. It combines furnace ammonia injection, known as selective non-catalytic reduction (SNCR), with LIDS to remove NO_x and to increase SO_2 removal and sorbent utilization. The calcium reaction path is the same as for LIDS. The ammonia reaction path begins with furnace injection of excess ammonia to control NO_x. Excess ammonia increases NO_x removal and inhibits ammonium bisulfate formation by kinetically favoring ammonium sulfate formation. Unreacted ammonia passes through the system to the dry scrubber where it reacts quantitatively with SO_2 in the cool, humid environment. This increases the overall SO_2 and reduces ammonia slip. Ammonia can be regenerated by treating the ash with a small quantity of water, as would be normally done to control dusting in an ash disposal system. In an alkaline environment, calcium displaces the ammonia in ammonia salts releasing ammonia gas.

The main objective of this phase was to demonstrate the synergistic aspects of this state-of-the-art technology including high NO_x removal, high SO_2 removal, ammonia regeneration, and

high calcium utilization. The goals for this phase were 60% NO_x removal, 95% SO_2 at Ca/S ratios less than 1.2, and greater than 90% ammonia regeneration efficiency.

Tests were run with an Ohio #6 coal containing 3.99% sulfur that produced about 3200 ppm of SO_2 at the scrubber inlet. The NO_x level at the burner outlet was maintained at about 630 ppm. Tests were conducted both with and without ammonia addition in the furnace. Furnace Ca/S ratios ranged from 1.00 to 1.48 and T_{as} from 10°F to 20°F. Furnace NH_3/NO_x ranged from 1.2 to 3.2. All tests were conducted with the addition of 1% $CaCl_2$ to the slurry solids to increase the moisture content of the baghouse ash to about 1.0%.

At a furnace Ca/S ratio of 1.0 and a scrubber T_{as} of 10°F, SO_2 removal increased from 87% to 99% when ammonia was used. This exceeds the SO_2 performance goals for this phase. A NO_x removal of 56% was attained at a furnace NH_3/NO_x ratio of 3.2. Ammonia slip was maintained below 10 ppm for this test and below 15 ppm for all tests by reducing the amount of excess ammonia introduced into the furnace. This shows that the performance goal of 60+% NO_x removal could be achieved easily by combining A+LIDS with Low NO_x burners that, alone, can reduce NO_x by 50%.

Magnesium-Based Dry Scrubbing

MgO-based scrubbing has been used commercially in wet processes but has never been used in a dry scrubber. The advantage of an MgO-based system is that the reaction product, magnesium sulfite ($MgSO_3$), decomposes at much lower temperatures than calcium sulfite ($CaSO_3$) and can, therefore, be economically regenerated. However, wet systems produce a sludge that must be dewatered and dehydrated before regeneration, which is a substantial energy penalty. Dry systems eliminate this step.

This process is similar to conventional lime dry scrubbing except that only small quantities of makeup MgO are required because the sorbent is regenerated. The process begins as dust-free flue gas enters the dry scrubber where it is contacted with atomized droplets of magnesium sulfite slurry. The droplets absorb SO_2 from the flue gas and dry before exiting the scrubber. Dried product is collected in a second particulate collection system, transported to a regeneration system, and heated to drive off SO_2 and produce new MgO for the scrubber. The gas from the regenerator contains about 15% SO_2 and can be used to produce sulfuric acid or other sulfur by-products.

The objective of this phase was to determine the feasibility of MgO-based dry scrubbing and fluidized bed sorbent regeneration. The performance goal was 90-95% SO_2 removal and 90% MgO recovery. The furnace was modified to burn natural gas to simulate dust-free flue gas. An SO_2 injection system was used to spike the flue gas to the desired SO_2 concentration.

Four tests were conducted at fresh Mg/S ratios ranging from 0.47 to 1.62 and T_{as} from 13 to 23°F. To simulate steady-state operation with recycle, $MgSO_3$ was added to the fresh slurry in the storage tank. Dry scrubber SO_2 removal ranged from 14 to 64% at Mg/S ratios of 0.47 to 1.63, which is about 15% lower than calcium-based systems. At a Mg/S ratio of 1.6 and a T_{as} of 15°F, the scrubber removed about 64% of the inlet SO_2 compared to 80% for a lime scrubber. However, no removal was observed across the baghouse causing the overall performance of this technology to be unacceptable. The regeneration process was not pursued due to the poor performance of the scrubbing system, and because it has already been reduced to commercial practice.

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Conclusions

SO_2 and NO_x Removal - The most significant accomplishment of this project is that the SO_2 removal goals were exceeded for all of the lime-based technologies. Only marginal performance was achieved with MgO-based scrubbing. This demonstrates the potential of B&W's advances to dry scrubbing, and shows the ability of these technologies to meet pollution control standards when burning high-sulfur, Ohio coal and other Eastern bituminous coals.

Results of Ca-Based Advanced Dry Scrubbing

Technology	Ca/S Ratio	T_{as}, °F	% SO_2 Removal
Dry Scrubbing	1.15	10	99
LIDS	1.44	11	99
A+LIDS - SO_2	1.00	10	99
- NO_x	3.2 NH_3/NO_x		56% NO_x

Operability at Low T_{as} - The success of the project can be attributed to two advances in dry- scrubbing technology; a new atomizer design, and a droplet impingement device (DID). These advances permitted trouble-free operation of the scrubber and baghouse at T_{as} of 10°F and less. Dry scrubber efficiency and sorbent utilization increase dramatically below T_{as} of 30°F. The proprietary B&W Durajet^{tm} atomizer minimizes deposition on the atomizer tip, that can lead to catastrophic deposition on scrubber walls and flues. The DID was installed at the scrubber outlet to prevent deposition downstream of the scrubber by selectively collecting any large, heavy, unevaporated slurry droplets present at low T_{as}. Smaller, lighter, dry particles follow the gas stream around the DID without being collected. Deposition on the DID was manageable at a T_{as} of 10°F as long as the feed slurry solids concentration was greater than about 10%. Inspections revealed no significant deposition on the atomizer, scrubber walls, flues, DID, or in the baghouse. Even at low T_{as}, the baghouse and downstream flues remained deposit-free and the ash from the baghouse flowed freely.

1.2   Economic Analysis and Commercialization

Within the last few decades wet scrubbing with lime or limestone slurries has become the dominant commercial FGD technology. Worldwide there are currently 581 FGD systems operating on a total capacity of about 150 GW_e[3]. Approximately 70% of the units, representing 124 GW_e of capacity, are based on lime or limestone wet scrubbing. About 20% of the units, or about 15 GW_e, utilize either sodium-based or lime-slurry dry scrubbing. The remaining 10% of the units use various regenerable processes of sorbent-injection technologies. Another 180 units, representing about 85 GW_e are planned or under construction, and will employ these technologies in approximately the same proportions. B&W will continue to develop and aggressively market two of the technologies studied during this project; Advanced Dry Scrubbing, and LIDS. Magnesium-based dry scrubbing will not be pursued because of the marginal performance demonstrated during the feasibility study. A+LIDS will be marketed as a niche technology when additional NO_x removal is needed on facilities seeking SO_2 control technology or as a means to polish SO_2 performance, while boosting NO_x removal, on future LIDS applications.

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Advanced Dry Scrubbing

The technological advancements in atomizer design and deposition control demonstrated during this project will be incorporated into the next generation of B&W dry scrubbers. However, dry scrubbing is perceived by the market to be applicable only to low-sulfur coal applications, and a fear is prevalent in the industry that they cannot be operated below a T_{as} of 30°F without catastrophic deposition. This fear developed from early bad experiences during dry scrubber start-ups, and little has been done since to achieve lower T_{as} at commercial scale. However, research continued to improve our understanding of basic dry scrubbing concepts, and great strides have been made in flow field modeling, atomizer design, and deposition control. This project has demonstrated that these advances can be used to improve dry scrubber SO_2 removal and sorbent utilization to levels previously attainable only by wet scrubbers. The next stage in the commercialization of these advances is the DOE Combustion 2000 Program discussed below.

LIDS and Combustion 2000

B&W is currently involved in the DOE Combustion 2000 Program with the purpose of developing an emissions control system capable of reducing SO_x emissions to one-third of that allowed under the current New Source Performance Standards (NSPS) and particulate emissions to one-half of the NSPS while addressing the concerns of solid waste generation and air toxics regulation. The work is being performed as an integral part of B&W's development of an advanced low-emission boiler system in a project entitled, "Engineering Development of Advanced Coal-Fired Low Emission Boiler Systems (LEBS)." The LEBS program comprises four distinct phases, with a total duration of 7 years (Fall 1992 - Fall 1999). B&W's first phase was completed in August 1994 and included the assessment of candidate subsystems and technologies, developing a research and development plan for the entire program, and preparing a preliminary design of a commercial generating unit. For control of SO_x and particulate emissions, numerous advanced flue gas cleanup options were assessed, and the B&W LIDS process was selected for the LEBS project. The second phase consists of engineering development, pilot-scale testing in B&W's 5 MBtu/hr SBS facility, and testing in B&W's new 100 MBtu/hr Clean Environment Development Facility (CEDF). The SBS tests are scheduled for January 1995 and the CEDF tests are scheduled for January 1996. Phase three entails the design of a 10-50 MW_{th} proof-of-concept test facility that will be built and tested in phase four.

Economic Analysis

Results from this project were used to estimate the operating costs of Advanced Dry Scrubbing, LIDS, and A+LIDS, and compare them to more conventional FGD technologies such as wet scrubbing and SCR. The analysis was based on the method described in the U.S. DOE Program Opportunity Notice DE-PSO1-88FE61530, and EPRI's Technical Assessment Guide (TAG)[4] Plant sizes of 100, 250, and 500 MW, and coal sulfur of 1.5, 2.5, and 3.5% were investigated. The calculations were based on a new plant design with a 30 year book life and 85% capacity factor. Operating costs were also estimated for T_{as} of 10°F and 30°F to illustrate the effect of low T_{as}.

The annual levelized costs for LIDS, Dry Scrubbing, and Wet FGD operating at 90% and 95% SO_2 removal, and for A+LIDS compared to a combined SCR/Dry FGD/FF are presented in $/ton of SO_2 removed and $/ton coal burned. For a 500 MW plant burning 3.5% S coal and achieving 90% SO_2 removal, operating at low T_{as} decreases costs from $322/ton of SO_2 to $259/ton for Dry FGD, and from $288/ton to $243/ton for LIDS. These costs are lower than for wet FGD at $272/ton of SO_2.

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Assuming 95% SO_2 removal, the costs are comparable to wet FGD at \$258/ton of SO_2 for wet FGD, \$262/ton for LIDS, and \$268/ton for Dry FGD. More importantly, however, this project demonstrated that it is reasonable to even consider the costs of dry FGD systems at SO_2 performance levels and high sorbent utilization previously reserved to wet scrubbers. Assuming 90% SO_2 and 60% NO_x removal for a 500 MW plant burning 3.5% S coal, A+LIDS costs less than the SCR/Dry FGD/FF system at \$287/ton and \$337/ton of SO_2 removed, respectively.

Waste disposal costs for Advanced Dry Scrubbing and LIDS were also estimated, by the method described above, for plants burning coals containing 1.5%, 2.5%, and 3.5% sulfur and scrubbing at 90% efficiency. For dry scrubbing, disposal costs ranged between \$45.50 and \$56.00 per ton of SO_2 removed at a T_{as} of 30°F for low- and high-sulfur coal respectively. By operating at a T_{as} of 10°F, disposal costs could be reduced by 15% and 25% to \$38.50 and \$42.00 per ton of SO_2 removed. Converting to \$/ton of coal burned gives \$1.52 and \$4.36 at a T_{as} of 30°F, and \$1.28 and \$3.27 at a T_{as} of 10°F for the same cases as above. For LIDS (and A+LIDS), disposal costs ranged between \$70.00 and \$87.50 per ton of SO_2 removed at a T_{as} of 30°F for low- and high-sulfur coal respectively. By operating at a T_{as} of 10°F, disposal costs could be reduced by 10% and 20% to \$63.00 and \$70.00 per ton of SO_2 removed. Converting to \$/ton of coal burned gives \$2.10 and \$5.44 at a T_{as} of 30°F, and \$2.33 and \$6.81 at a T_{as} of 10°F for the same cases.

Annual Levelized Costs @ 90% SO_2 Removal

\$/ton SO_2 Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=30°F	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	565	533	597	564	895
100	2.5	443	395	443	403	658
100	3.5	390	327	377	332	494
250	1.5	446	414	437	404	640
250	2.5	371	323	349	309	421
250	3.5	339	276	311	266	327
500	1.5	406	374	383	350	504
500	2.5	347	299	316	276	337
500	3.5	322	259	288	243	272

\$/ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=30°F	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	15.3	14.4	16.1	15.2	24.2
100	2.5	19.9	17.8	19.9	18.1	29.6
100	3.5	24.6	20.6	23.8	20.9	31.1
250	1.5	12.0	11.2	11.8	10.9	17.3
250	2.5	16.7	14.5	15.7	13.9	18.9
250	3.5	21.4	17.4	19.6	16.8	20.6
500	1.5	11.0	10.1	10.3	9.5	13.6
500	2.5	15.6	13.5	14.2	12.4	15.2
500	3.5	20.3	16.3	18.1	15.3	17.1

Basis: 90 % SO_2 Removal

New Plant Design

30 yr book life, 85% Capacity Factor

Annual Levelized Costs @ 95% SO₂ Removal

$/ton SO₂ Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD $T_{as}=10°F$	LIDS $T_{as}=30°F$	LIDS $T_{as}=10°F$	Wet FGD
100	1.5	527	596	561	855
100	2.5	397	450	411	631
100	3.5	332	388	346	474
250	1.5	415	445	387	616
250	2.5	329	361	322	406
250	3.5	284	325	284	316
500	1.5	377	393	359	486
500	2.5	306	330	291	326
500	3.5	268	303	262	258

30 yr book life, 85% Capacity Factor

$/ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD $T_{as}=10°F$	LIDS $T_{as}=30°F$	LIDS $T_{as}=10°F$	Wet FGD
100	1.5	15.0	17.0	16.0	24.4
100	2.5	18.9	21.4	19.5	30.0
100	3.5	22.1	25.8	23.0	31.5
250	1.5	11.8	12.7	11.0	17.6
250	2.5	15.6	17.1	15.3	19.3
250	3.5	18.9	21.6	18.9	21.0
500	1.5	10.7	11.2	10.2	13.9
500	2.5	14.5	15.7	13.8	15.5
500	3.5	17.8	20.1	17.4	17.2

Basis: 95 % SO₂ Removal

New Plant Design

30 yr book life, 85% Capacity Factor

Annual Levelized Costs @ 90% SO₂, 60% NO$_x$ Removal

$/Ton SO₂ + NO$_x$ Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD, SCR & FF	A+LIDS $T_{as}=10°F$
100	3.5	405	371
250	3.5	354	309
500	3.5	337	287

$/Ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD, SCR & FF	A+LIDS $T_{as}=10°F$
100	3.5	26	23
250	3.5	22	19
500	3.5	21	18

Basis: 90 % SO₂ Removal

60 % NO$_x$ Removal

New Plant Design

30 yr book life, 85% Capacity Factor

2.0 FULL REPORT

2.1 INTRODUCTION

Dry Scrubbing is a common commercial process that has been limited to low- and medium-sulfur coal applications because high-sulfur coal requires more reagent than can be efficiently injected into the process. Babcock & Wilcox has made several advances that extend dry scrubbing technologies to higher sulfur coals by allowing deposit-free operation at low scrubber exit temperatures. This not only increases the amount of reagent that can be injected into the scrubber, but also increases SO_2 removal efficiency and sorbent utilization.

The objectives of this project were to demonstrate, at pilot scale, that advanced, dry-scrubbing-based technologies can attain the performance levels specified by the 1990 Clean Air Act Amendments for SO_2 and NO_x emissions while burning high-sulfur coal, and that these technologies are economically competitive with wet scrubber systems. The use of these technologies by utilities in and around Ohio, on new or retrofit applications, will ensure the future of markets for high-sulfur coal by creating cost-effective options to coal switching.

The project included four test phases; Phase 4 - Advanced Dry Scrubbing, Phase 5 - LIDS, Phase 6 - A+LIDS, and Phase 7 - Regenerable MgO Dry Scrubbing. Phase 4 was divided into two sub-phases; 4A - testing with medium- to low-sulfur coal, and 4B - testing with high-sulfur coal. Each phase represented a different, advanced dry scrubbing technology in various stages of development. The final test phase was completed in August. A separate Milestone Report was issued for each test phase and the results are presented below.

2.1.1 Changes to the Statement of Work (SOW) and Milestone Plan (MP)

Because of perceived facility schedules at the time of writing, the SOW names Phase 4 as Dry Scrubber Testing and Phase 5 as LIDS, while the MP has the reverse. As it turned out, the SOW had the correct names for the phases. Also, Phase 4 was divided into two sub-phases to accommodate facility schedules. The sub-phases, named 4A and 4B represented tests with low- and high-sulfur coal, respectively. The SOW outlined a range of test conditions but gave no specific test matrix. More specific test matrices are contained in the MP along with the disclaimer that they were subject to change based on the results of completed tests and unforeseen problems, and that OCDO would be notified of changes (SOW p. 5). This occurred during Phase 5 because of the poor performance attained during the initial tests, and during Phase 7 because of the poor overall performance. The changes are summarized below.

Phase 4A - Advanced Dry Scrubbing with Low-Sulfur Coal

Eight tests were originally planned for this sub-phase at a range of Ca/S and T_{as}. Nine tests were completed, and all goals were met or exceeded.

Phase 4B - Advanced Dry Scrubbing with High-Sulfur Coal

Because of facility schedules, this sub-phase was completed after Phase 5. Also, because of problems encountered during Phase 5, the test matrix shortened from 14 tests to 10 tests. This change was approved by OCDO in a letter dated 4/28/94. Ten tests were completed, and all goals were met or exceeded.

Phase 5 - LIDS

Originally, the goal of this test was simply to extend the data base for this technology, established in a 1991 test co-sponsored by OCDO, to lower T_{as} and higher SO_2 removal made possible by technology advances. However, initial tests were unable to duplicate 1991 results, and additional tests were conducted to determine the cause of the differences. To help fund these tests an additional $35,000 was contributed by B&W. The early problems with this phase necessitated a reorganization of the remaining schedule. OCDO approved the changes to Phases 5, 4B, 6, and 7 in a letter dated April 28, 1994.

Five tests were originally planned for this phase, but because of the problem mentioned above, 31 LIDS tests were completed, including four in Phase 6. A solution to the problem was found and, eventually, all goals were met or exceeded

Phase 6 - A+LIDS

As mentioned above, the matrix for this phase was reduced from 8 tests to 4 tests. The change was approved by OCDO in the letter. In all, 13 tests were completed that included baseline tests and four additional LIDS tests at the same conditions but without ammonia injection. All goals were met or exceeded.

Phase 7 - Regenerable MgO-Based Dry Scrubber

The main objective of this phase was to investigate the feasibility of this untried technology. Originally, three tests were planned. This was later reduced to one test and approved by OCDO in the 4/28/94 letter. The number of tests was later increased to five, because of the savings possible by conducting the regeneration portion of the test, a proven commercial process, at bench-scale instead of pilot-scale. However, more time was required to complete the scrubber tests because of unforeseen problems with slurry handling. Since the scrubbing system only achieved marginal performance, the technology was deemed not feasible at this stage of development, and the subsequent regeneration tests were not pursued. OCDO was notified of these changes, and approved them on 12/20/94.

2.2   TECHNICAL DISCUSSION

All tests were conducted in B&W's pilot-scale furnace known as the Small Boiler Simulator (SBS). The SBS is PC-fired and rated at 5 million Btu/hr. Other test equipment included a 5 ft. diameter, vertical, down-flow dry scrubber with dual fluid atomization, and a pulse-jet baghouse. Appendix A is a detailed facility description. A typical test series included start-up, a test period, and shutdown. Start-up involved operating the furnace for several hours to reach thermal equilibrium. Equilibrium was determined by monitoring several TC's and TC grids at various locations in the gas path. Start-up also included subsystem shakedown, analyzer calibration and slurry preparation. For phases that involved furnace sorbent injection, start-up was followed by collecting baseline data to determine background furnace SO_2 concentration before sorbent injection. This value was later used to calculate furnace SO_2 removal. This procedure was necessary because high temperatures make it difficult to sample flue gas ahead of sorbent injection. A typical test period lasted between 3 and 12 hours depending on how fast steady state was attained. Tests with recycle required longer periods to reach steady state. Two 55 gallon drums were used to make slurry from waste ash. A portion of the ash from the baghouse hoppers was metered to a drum and mixed with water. The resulting slurry

was given one hour to slake before being transferred to a storage tank. Steady-state was determined by monitoring analyzer strip charts, and by hourly titrations of slurry samples for sorbent concentration. After attaining steady-state, the analyzer banks were recalibrated and a data set spanning five minutes was collected and stored. Coal, slurry, and ash samples were collected after each data set.

System performance was evaluated by measuring the SO_2, O_2, and NO_x concentration at the furnace, scrubber and baghouse exits. Removal efficiency for each subsystem was determined by correcting the SO_2 and NO_x values to 3% O_2. System operability was determined by monitoring deposition and pressure drop in the convective pass, scrubber, baghouse and DID. Chemical analyses were performed on coal, slurry, and ash samples from each test. Analyses included coal ultimate analysis, slurry reactivity and solids concentration, and an elemental and species analysis of the waste material. A dust loading at the stack was performed to test the collection efficiency of the baghouse. The project included four tests phases, each representing a different dry-scrubbing-based technology. The technologies and test results are described below.

2.2.1   Phase 4 - Advanced Dry Scrubbing

Figure 1 is a schematic of B&W's dry scrubbing process that begins by slaking pebble lime (CaO) to produce a highly reactive slurry. The slurry is atomized with high-pressure air and injected into the dry scrubber to remove SO_2. The scrubber is carefully designed so that the slurry droplets dry before reaching the exit or impinging on the walls. Unreacted sorbent, flyash, and reaction products are collected on the baghouse bags. Additional SO_2 removal occurs in the baghouse filter cake because of the extended contact time. A portion of the baghouse waste material can be mixed with fresh sorbent to increase sorbent utilization or to increase the solids concentration of the slurry to improve drying characteristics.

The important parameters in this process are the amount of lime slurry injected into the scrubber relative to the amount of SO_2 in the flue gas, known as the stoichiometric ratio (SR), or moles of $Ca(OH)_2$ per mole of SO_2 (Ca/S), and the approach to saturation temperature (T_{as}) of the scrubber and baghouse. T_{as} is the difference between the flue gas wet-bulb and dry-bulb temperatures and denotes the degree of saturation of the flue gas. A T_{as} of 0°F means the flue gas is completely saturated. A T_{as} of 20°F is equivalent to about 50% relative humidity. Scrubber and baghouse SO_2 removal efficiency greatly increase as T_{as} approaches zero. However, problems with deposition and ash handling also increase as T_{as} decreases. Typical utility dry scrubbers operate at T_{as} greater than 30°F to avoid slurry deposits on the scrubber walls and baghouse. B&W's dry scrubbing technology advancements allow deposit-free operation at a T_{as} of 10°F.

The major chemical reactions that occur in the slaker, dry scrubber, and baghouse are shown below. For greatest efficiency, the reactions must take place in the liquid phase. Once the slurry droplet has dried, the reactions no longer proceed.

2-3

FIGURE 1 Dry Scrubbing Process Schematic

Advanced Dry Scrubbing

Slaking Reaction:
$$CaO + H_2O \rightarrow Ca(OH)_2 + Heat$$

Scrubber and Baghouse Reactions:
$$Ca(OH)_2 \rightarrow Ca^{+2} + 2OH^-$$
$$SO_{2(aq)} + H_2O \rightarrow HSO_3^- + H^+$$
$$SO_{2(aq)} + OH^- \rightarrow HSO_3^-$$
$$HSO_3^- + OH^- \rightarrow SO_3^{-2} + H_2O$$
$$Ca^{+2} + SO_3^{-2} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O$$

$$\overline{Ca(OH)_2 + SO_2 \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O + \tfrac{1}{2}H_2O}$$
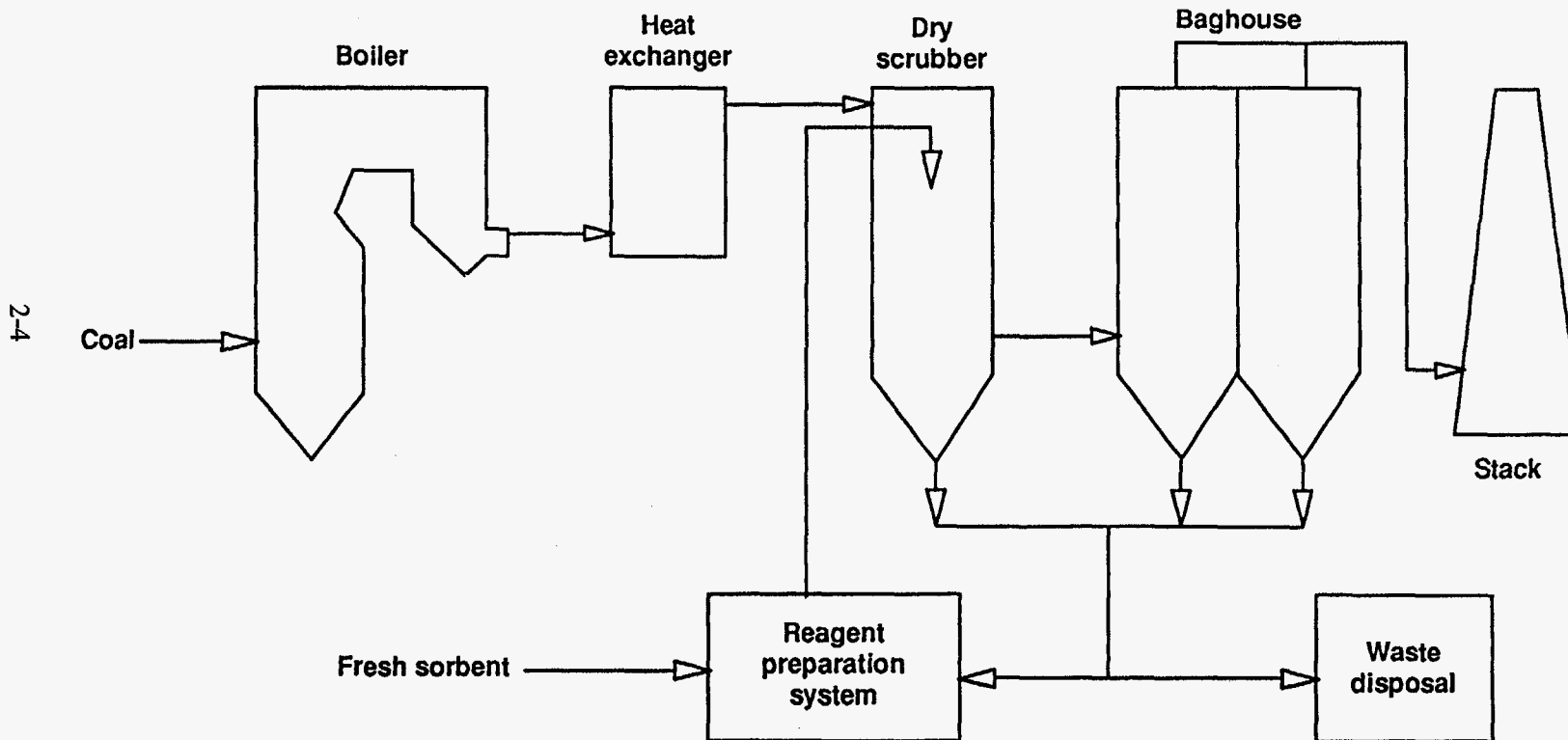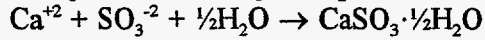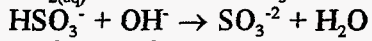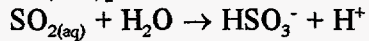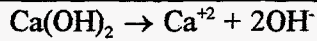
2.2.2   Phase 4 - Results

The objectives of this phase were to establish a sound database for comparison to wet scrubbing and other dry-scrubbing technologies, and to demonstrate B&W's advances in dry scrubbing. The performance goal was 90-95% SO_2 removal at a Ca/S ratio less than 1.5 using a high-sulfur Ohio Coal.

Advanced dry scrubbing concepts were tested in two sub-phases. Phase 4A was conducted with a washed Ohio #6 coal having an average sulfur content of 1.32% by weight. It produced about 1000 ppm of SO_2 at the scrubber inlet. The coal for Phase 4B was an unwashed Ohio #6 with an average sulfur of 4.35%. This coal produced about 3600 ppm of SO_2 at the inlet of the scrubber. Table 1 shows typical analyses for these coals.

TABLE 1   Typical Coal Analyses - Phase 4

	Phase 4A		Phase 4B	
	As Fired	Dry	As Fired	Dry
Gross Heating Value				
Btu/lb.	12005	13048	12377	12794
Btu/lb. (M&A Free)		14124		14443
Ultimate Analysis, Wt. %				
Moisture	7.99		3.26	
Carbon	68.44	74.38	69.13	71.46
Hydrogen	4.80	5.22	4.84	5.00
Nitrogen	1.14	1.24	1.30	1.34
Sulfur	1.32	1.43	4.17	4.31
Ash	7.01	7.62	11.05	11.42
Oxygen (by difference)	9.30	10.11	6.25	6.47

Tables 2 and 3 are summaries of important operating parameters and performance results. Table 2 represents Phase 4A and Table 3 represents 4B. The upper portion of the tables contain flue gas temperatures and pressure drop for the major components, and atomizer slurry and air flows. The lower portion contains the fresh and total (fresh plus recycle) scrubber stoichiometry, approach to saturation temperature (T_{as}), SO_2 concentrations adjusted to 3% oxygen, SO_2 removal data, sorbent utilization, and the flue gas flow rate at the scrubber inlet. A list of nomenclature is provided with each table.

Nineteen tests were conducted during Phase 4 encompassing fresh Ca/S ratios ranging from 0.62 to 1.91 and T_{as} from 8°F to 30°F. Several tests in Phase 4B were conducted at high excess air to dilute the SO_2 concentration in the flue gas. This was necessary because the coal delivered for this phase had a much higher sulfur content (4.3%) than planned (3.5%). Oxygen concentration at the scrubber inlet increased from 3.0% to 5.0%, but was still in the range of a typical commercial unit. The SO_2 concentration decreased from 3700 ppm to 3100 ppm which is equivalent to a decrease in coal sulfur from 4.3% to 3.7% correcting to 3% oxygen.

Dry Scrubber Performance

Figure 2 shows dry scrubber SO_2 removal as a function of fresh Ca/S ratio. Removal is based on the scrubber inlet and outlet SO_2 concentration corrected to 3% oxygen. Data is grouped by tests at similar T_{as}. Solid data points represent tests utilizing recycle. The dashed line dissects the graph between the high- and low-sulfur coal tests. The high-sulfur coal used for this test limited the Ca/S ratio to a maximum 1.15. The figure shows that at similar Ca/S ratios, lowering the T_{as} by 10°F increased SO_2 removal by about 10%. The use of recycle had little effect during tests with high-sulfur coal because little recycle could be used. Tests with low-sulfur coal showed a greater effect of recycle. At a T_{as} of 10°F, SO_2 removal increased by 10% with the use of recycle. The best results were attained at a fresh Ca/S ratio of 1.0 and a T_{as} of 10°F, where the scrubber achieved 61% SO_2 with high-sulfur coal, and 78% with low-sulfur coal.

Baghouse Performance

Figure 3 shows baghouse SO_2 removal as a function of fresh Ca/S ratio for both coal types. Removal is based on the baghouse inlet and outlet SO_2 concentration corrected to 3% oxygen. Data is grouped by tests representing similar T_{as}. Solid data points represent tests utilizing recycle. The dashed line divides the high- and low-sulfur tests. The figure shows that T_{as} had a greater effect on baghouse SO_2 performance than on scrubber performance. At a Ca/S of 1.0, performance increased from about 44% to 72% when T_{as} was lowered from 20°F to 10°F. This is likely due to the free moisture in the baghouse ash that increased from 0.7% to 1.6% for the same conditions. The reactions require liquid water, so the efficiency increased when more water was present. These tests showed that ash moisture must be above 1% to achieve high SO_2 removal efficiency. It should be noted that at 1.6% moisture, the ash appeared dry and caused no handling problems. Also, the use of recycle had little affect on baghouse SO_2 removal.

Overall, the baghouse is a very efficient pollution control device at low T_{as} (or high ash moisture). Tests at fresh Ca/S ratios above 1.2 and a T_{as} of 10°F achieved over 90% SO_2 removal. However, there is a fear prevalent in industry that baghouses cannot operate below a T_{as} of 30°F without forming hard deposits on the bags. This phase demonstrated that the use of Teflon coated bags and B&W's new Droplet Impingement Device (DID) can prevent deposition in the baghouse. A post-inspection showed only a slight (1/16 in.), dry, fluffy deposit on the bags after weeks of operation below T_{as} of 30°F, and to 10°F for 48-hour periods.

TABLE 2 Summary of Phase 4A

Test	Temperature (°F)			Differential Pressure (in.)			Atomizer Streams						Atom. Air	Vent Air	A:W
							Slurry	%	Other	%	Total	%	Air	Air	
	DSin	DSout	Twb	HX	DS	BH	(lb/hr)	Solids	(lb/hr)	Solids	(lb/hr)	Solids	(lb/hr)	(lb/hr)	(lb/lb)
1	297	151	122	11.3	1.2	4.2	159	12.8	20	0.0	180	11.3	50	36	0.28
2	299	142	122	11.2	1.3	4.4	164	12.7	45	0.0	208	10.0	55	48	0.26
3	299	128	120	11.4	1.3	3.7	152	13.0	76	0.0	228	8.6	56	55	0.24
4	298	153	125	11.6	1.2	4.0	105	12.7	65	0.0	170	7.8	49	44	0.29
5	298	144	124	11.7	1.3	3.3	102	12.7	79	0.0	180	7.2	50	44	0.28
6	298	133	124	11.3	1.3	4.0	103	13.7	102	0.0	205	6.9	54	55	0.26
7	299	132	123	11.4	1.2	4.2	109	16.4	174	44.1	282	33.4	79	69	0.28
8	300	144	124	10.0	1.2	3.9	124	17.1	114	45.4	239	30.6	61	58	0.25
9	300	152	123	10.6	1.3	2.9	94	17.1	139	46.4	232	34.6	61	57	0.26
10	300	133	122	10.6	1.2	3.6	83	16.3	209	44.9	292	36.8	78	69	0.27

Test	SR		T_{as}	SO$_2$ Data (ppm @ 3% O$_2$, dry)			SO$_2$ Removal (%)			Ca	DSin Gas
	Fresh	Total	(°F)	DSin	DSout	BHout	DS	BH	Total	Util.	(lb/hr)
1	1.63	1.63	28	983	365	158	62.9	56.6	83.9	51.4	4913
2	1.69	1.69	20	991	263	11	73.5	95.9	98.9	58.5	4970
3	1.74	1.74	8	922	78	2	91.6	97.8	99.8	57.4	4900
4	1.17	1.17	28	952	428	284	55.0	33.6	70.1	60.2	4872
5	1.10	1.10	19	952	357	129	62.5	64.0	86.5	78.4	4919
6	1.21	1.21	10	913	232	22	74.6	90.4	97.6	80.3	4885
7	1.35	1.35	9	1087	58	5	94.7	91.1	99.5	73.7	4917
8	1.91	2.01	20	918	140	1	84.7	99.6	99.9	52.2	4910
9	1.16	1.32	30	1119	414	219	63.0	47.2	80.5	69.4	4955
10	1.16	1.16	10	987	143	5	85.5	96.4	99.5	85.7	4841

Nomenclature

Twb = Wet Bulb Temperature

HX = Heat Exchanger

DS = Dry Scrubber

BH = Baghouse

Other = Recycle Slurry or Make-up Water

A:W = Air to Liquid Ratio of Atomizer

SR = Calcium to Sulfur Molar Ratio

T_{as} = Approach to Saturation Temperature

Util = Sorbent Utilization = SO$_2$ Removal/SR

TABLE 3   Summary of Phase 4B

Test	Temperature (°F)			Differential Pressure (in.)			Atomizer Streams						Atom Air	Vent Air	A:W
	DSin	DSout	Twb	HX	DS	BH	Slurry (lb/hr)	% Solids	Other (lb/hr)	% Solids	Total (lb/hr)	% Solids	(lb/hr)	(lb/hr)	(lb/lb)
1	301	143	123	9.6	1.3	4.7	202	17.5	0	0.0	202	17.5	73	77	0.36
2	298	132	122	9.9	1.3	4.9	198	17.6	19	0.0	218	16.0	76	85	0.35
3	303	133	123	10.8	1.3	4.5	228	17.5	0	0.0	228	17.5	75	85	0.33
4	296	142	123	9.9	1.4	3.8	203	17.7	0	0.0	203	17.7	75	86	0.37
5	300	130	121	17.0	1.7	5.3	263	17.5	0	0.0	263	17.5	82	101	0.31
6	301	140	121	16.2	1.7	7.4	233	17.9	0	0.0	233	17.9	63	86	0.27
7	301	131	120	17.4	1.7	7.1	229	17.8	30	38.6	259	20.2	70	91	0.27
8	305	130	120	14.5	1.6	8.0	230	18.1	21	39.5	251	19.9	71	90	0.28
9	303	130	120	14.9	1.6	7.8	217	12.0	34	41.8	251	16.0	73	90	0.29

Test	SR		T_{as}	SO$_2$ Data			SO$_2$ Removal (%)			Ca	DSin Gas
	Fresh	Total by ICP	(°F)	DSin	DSout	BHout	DS	BH	Total	Util.	(lb/hr)
				(ppm @ 3% O$_2$, dry)							
1	0.80	0.80	20.5	3671	2158	1751	41.2	18.9	52.3	65.8	4733
2	0.75	0.75	9.5	3826	1931	1032	49.5	46.6	73.0	96.8	4735
3	0.87	0.87	9.8	3719	1703	696	54.2	59.1	81.3	93.0	4821
4	0.79	0.79	19.1	3723	2147	1631	42.3	24.0	56.2	70.8	4764
5	1.15	1.15	9.7	3458	1301	111	62.4	91.5	96.8	84.3	5449
6	0.99	0.99	19.4	3573	1750	1261	51.0	27.9	64.7	65.2	5366
7	0.93	0.95	10.6	3677	1521	543	58.6	64.3	85.2	91.4	5488
8	0.97	0.98	9.6	3583	1420	519	60.4	63.5	85.5	88.5	5406
9	0.62	0.63	9.8	3475	1679	1275	51.7	24.0	63.3	102.0	5432

Note: 102% Utilization is within the limits of error of measurements.

Nomenclature

Twb = Wet Bulb Temperature

HX = Heat Exchanger

DS = Dry Scrubber

BH = Baghouse

Other = Recycle Slurry or Make-up Water

A:W = Air to Liquid Ratio of Atomizer

SR = Calcium to Sulfur Molar Ratio

ICP = Analysis by Inductivrly Coupled Plasma

T_{as} = Approach to Saturation Temperature

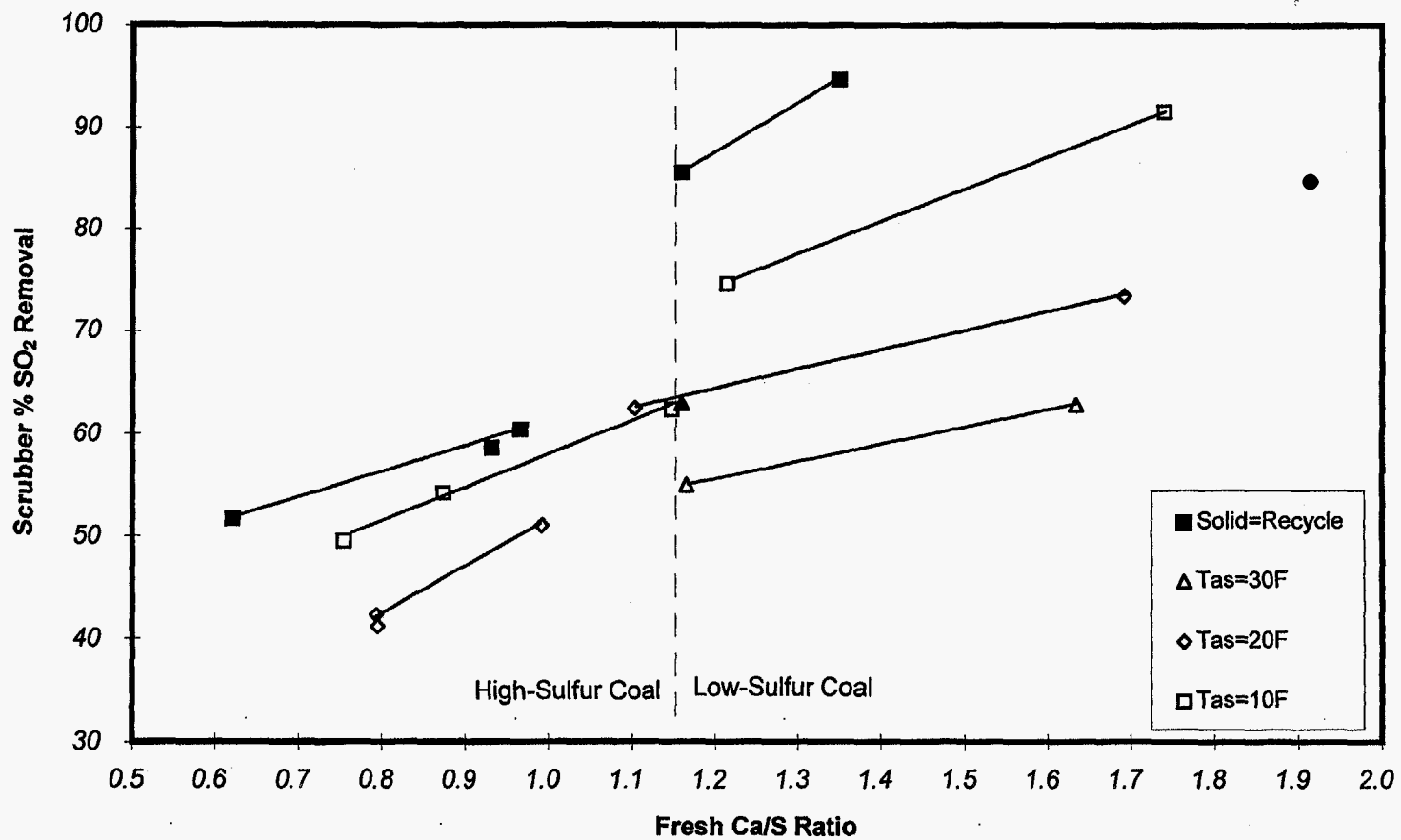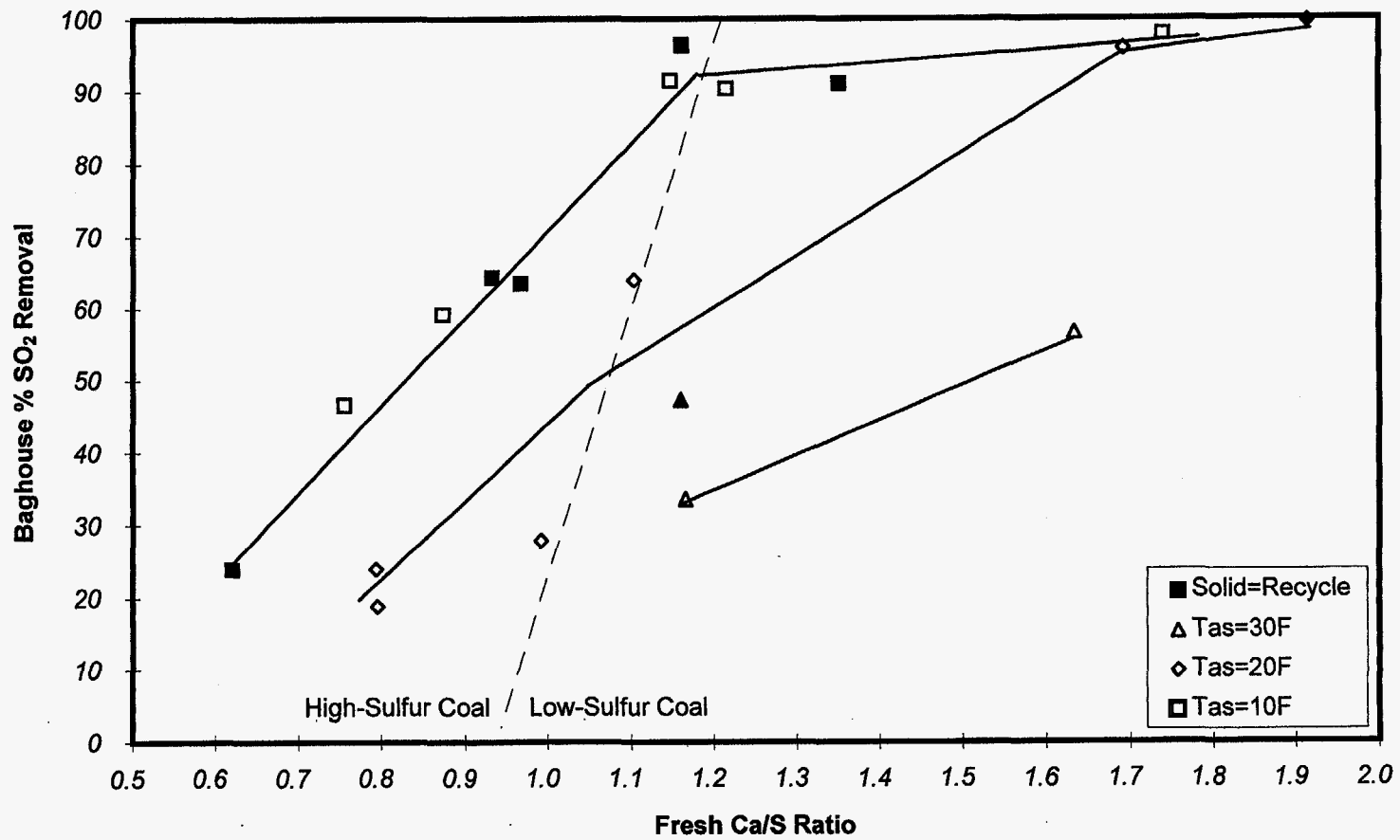Util = Sorbent Utilization = SO$_2$ Removal/SR

FIGURE 2   Phase 4 - Dry Scrubber SO$_2$ Removal

FIGURE 3   Phase 4 - Baghouse SO_2 Removal

Total SO$_2$ Removal

Figure 4 shows total SO$_2$ removal as a function of fresh Ca/S ratio for both coal types. Data is grouped by tests with similar T$_{as}$. Solid data points represent tests utilizing recycle. Again, the dashed line divides the high- and low-sulfur tests. The figure shows that at T$_{as}$ of 10°F there was no difference in performance due to coal sulfur, only in the maximum Ca/S ratio that could be attained. The curves representing high- and low-sulfur tests at a T$_{as}$ of 10°F meet at a Ca/S ratio of about 1.1.

Performance goals were exceeded for this phase. At a Ca/S ratio of 1.15 and a T$_{as}$ of 10°F, over 99% SO$_2$ removal was attained. Calcium utilization for several tests was above 90%. This demonstrates that dry scrubbing can achieve the same SO$_2$ performance as wet scrubbers at the pilot scale when operated at low T$_{as}$, and that low T$_{as}$ can be maintained without deposition by utilizing the latest advancements.

Deposition Control by DID

All tests were conducted with the B&W patented Droplet Impingement Device (DID) installed at the scrubber outlet. Figure 5 is a schematic of the DID as it was used in the pilot scrubber. The DID consists of two staggered rows of anodized aluminum angle that extend across the entire cross section of the scrubber like a barbecue grill. Its purpose is to prevent deposition downstream of the scrubber by selectively collecting large, unevaporated slurry droplets, that may be present at low T$_{as}$. Smaller, lighter and dry particles have much less momentum and follow the gas stream around the DID without being collected. Provisions were made to clean the DID by rapping, rotating, or both.

The DID performed well during all test phases. A significant finding was that the DID was only needed when the scrubber was operated at very low T$_{as}$ with slurries containing less then 10% solids. Under these conditions, incomplete evaporation and low initial solids concentrations resulted in large, wet, slurry droplets at the scrubber exit. These droplets impinged the DID and gradually built deposits that had to be removed by rotating and rapping the DID. In a few cases, the deposits were moist and difficult to remove. However, even for these cases, the downstream ducts and equipment remained deposit-free demonstrating that the DID concept is viable during extreme or upset conditions. In tests using slurries above 10% solids, little material collected on the DID at any condition.

Post-test inspections revealed no significant deposition on the atomizer, scrubber walls, flues, DID, or in the baghouse. Even at low T$_{as}$, the baghouse and downstream flues remained deposit-free and the ash from the baghouse flowed freely. The pilot facility was able to operate at much lower T$_{as}$ than commercial dry scrubbers due to advances in atomizer design and the use of the DID.

2.2.3 Phase 5 - LIDS

LIDS, Limestone Injection Dry Scrubbing, combines furnace limestone injection and dry scrubbing to achieve high levels of SO$_2$ removal and reduce reagent costs while burning high-sulfur coal. A schematic of the LIDS process is shown in Figure 6. In the process, pulverized limestone is injected into the upper furnace and reacts with a portion of the SO$_2$. Unreacted lime and waste material pass through the system and are collected in a baghouse. A portion of this material is slaked to produce a highly-reactive slurry for dry scrubbing. The slurry is atomized with high-pressure air and injected into the dry scrubber to remove more SO$_2$. Additional SO$_2$ removal occurs in the

FIGURE 4 Phase 4 - Total SO₂ Removal
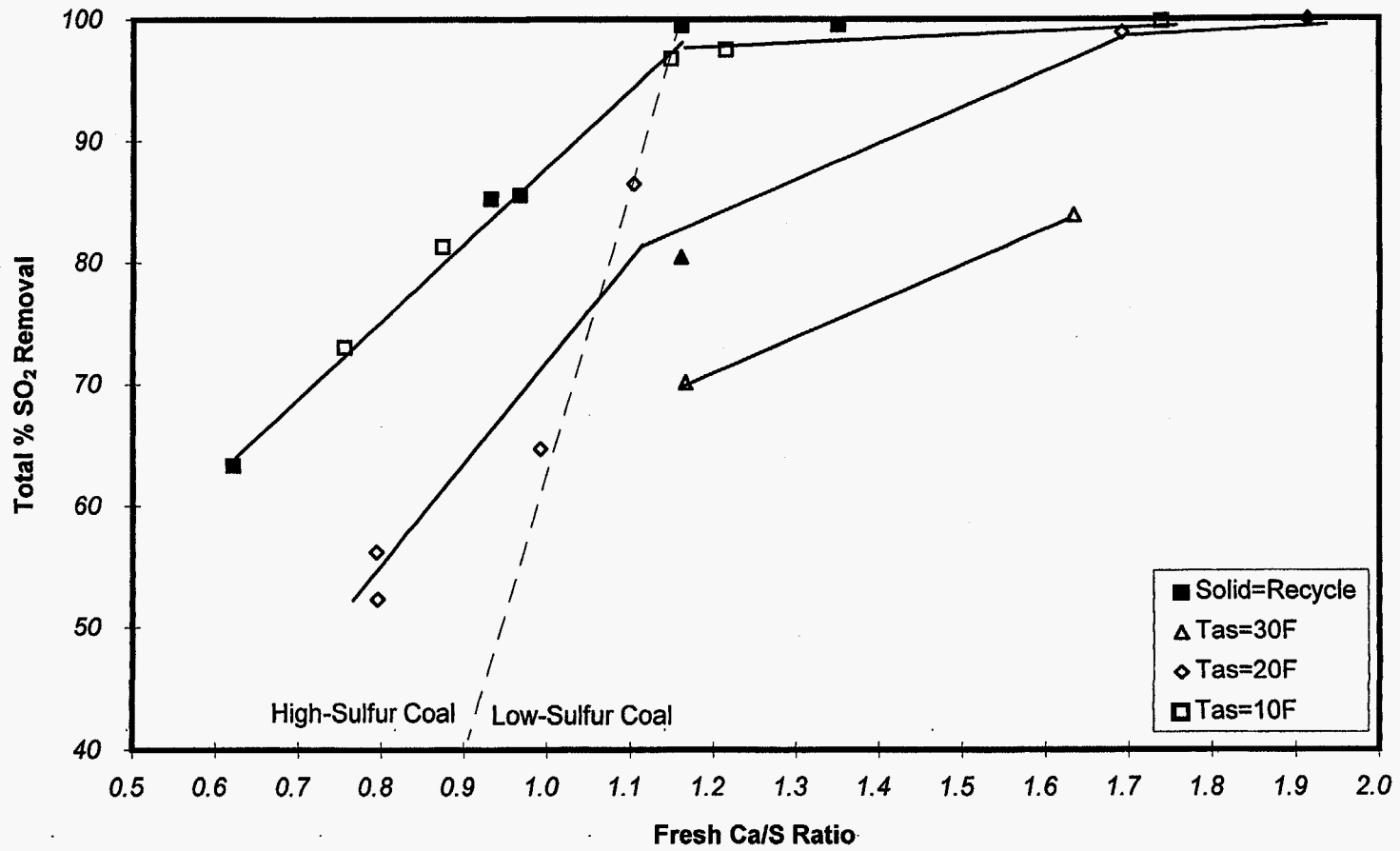
FIGURE 5 Droplet Impingement Device (DID) in the Pilot Scrubber
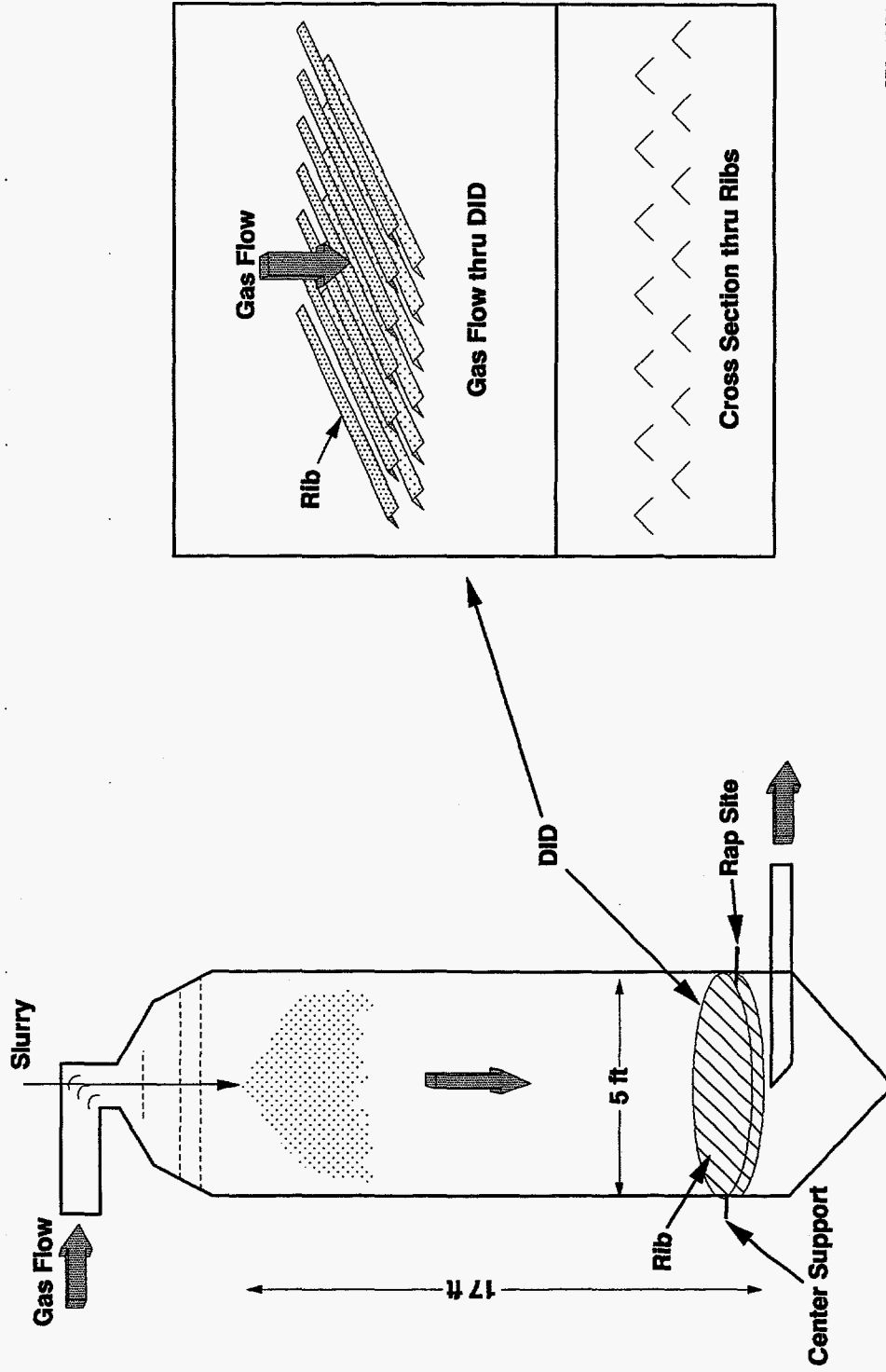
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FIGURE 6 LIDS Process Schematic

LIDS
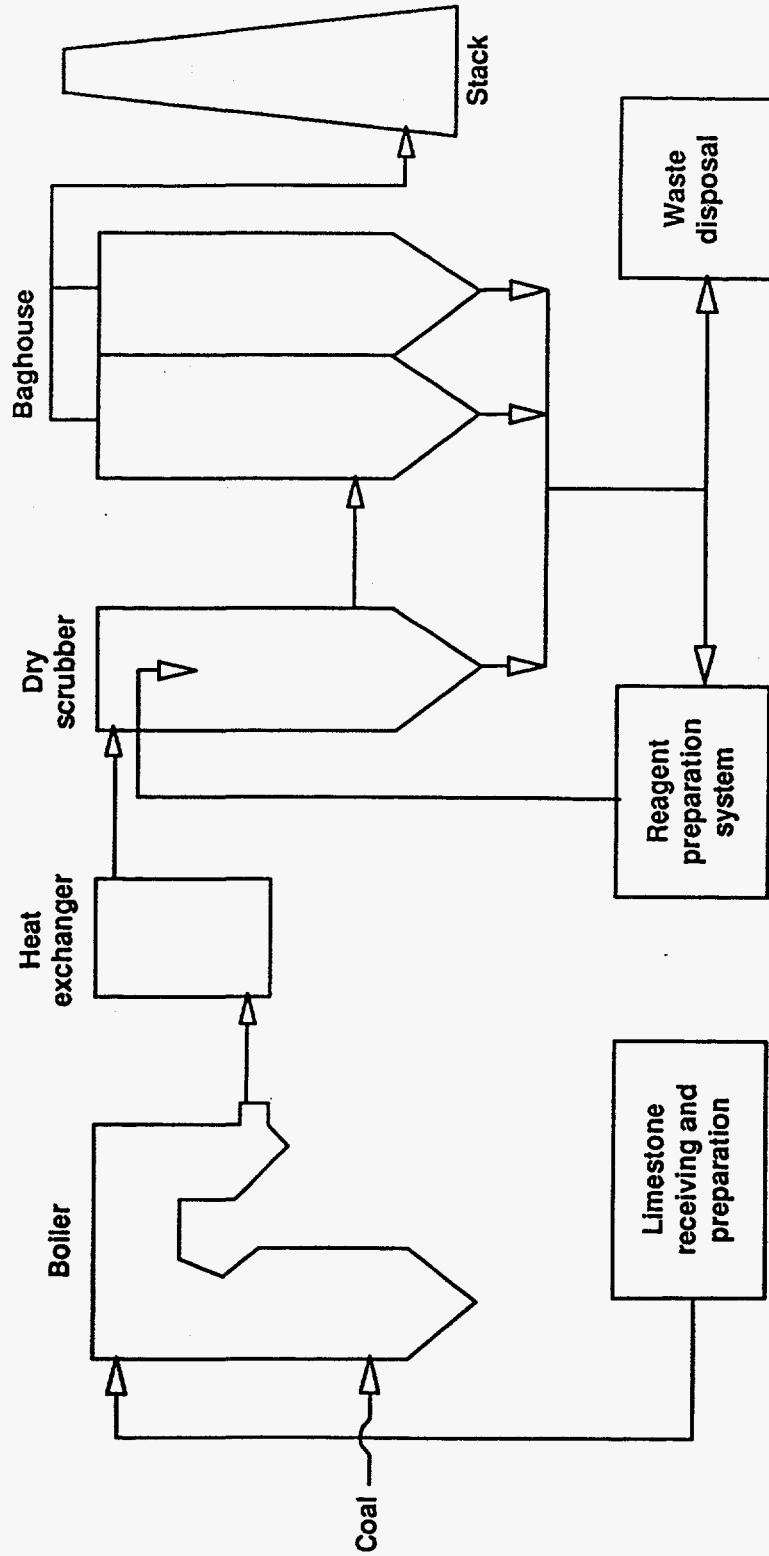
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baghouse. The differences between LIDS and dry scrubbing are that SO_2 removal occurs in **three** stages (furnace, scrubber, baghouse), and that the waste ash from limestone furnace injection is the only source of lime for the dry scrubber, no pebble lime is needed.

The important parameters in this process are the amount of reagent injected in the furnace (SR or Ca/S), the temperature at the injection zone, and the T_{as} of the scrubber and baghouse. The optimum temperature range for limestone injection is 2100-2300°F. Injection at higher temperatures causes limestone dead burning which decreases sorbent reactivity. Injection at lower temperatures inhibits calcination which also reduces sorbent reactivity. As in dry scrubbing, scrubber and baghouse SO_2 removal efficiency increase as T_{as} decreases, as does potential problems with deposition and ash handling.

The major chemical reactions that occur in the furnace, slaker, scrubber, and baghouse are shown below. For greatest efficiency, the scrubber and baghouse reactions must take place in the liquid phase. Once the slurry droplet has dried, the reactions no longer proceed.

<u>In-Furnace Reactions:</u>
$$CaCO_3 + Heat \rightarrow CaO + CO_2$$
$$CaO + SO_2 + \tfrac{1}{2}O_2 \rightarrow CaSO_4$$

<u>Slaking Reaction:</u>
$$CaO + H_2O \rightarrow Ca(OH)_2 + Heat$$

<u>Scrubber and Baghouse Reactions:</u>
$$Ca(OH)_2 \rightarrow Ca^{+2} + 2OH^-$$
$$SO_{2(aq)} + H_2O \rightarrow HSO_3^- + H^+$$
$$SO_{2(aq)} + OH^- \rightarrow HSO_3^-$$
$$HSO_3^- + OH^- \rightarrow SO_3^{-2} + H_2O$$
$$Ca^{+2} + SO_3^{-2} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O$$

$$\overline{Ca(OH)_2 + SO_2 \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O + \tfrac{1}{2}H_2O}$$
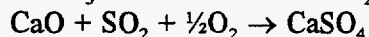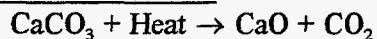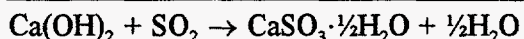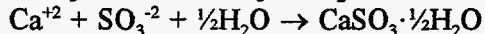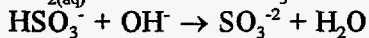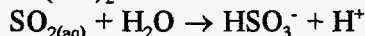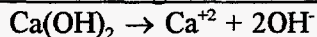
2.2.4 Phase 5 - Results

The objective of this phase was to build on the strong data base attained during the OCDO sponsored Pilot-Scale LIDS Demonstration in 1990-91[1,2]. Knowledge gained from previous work, along with recent dry scrubbing advances, were used to extend this technology to higher levels of SO_2 removal at lower Ca/S ratios. This phase tested lower T_{as}, higher furnace injection temperatures, higher slurry solids, and continuous operation. Tests in 1991 showed that 92% SO_2 removal was possible at a Ca/S ratio of 1.9. The goal of this phase was to demonstrate that LIDS is capable of 95% SO_2 removal at a Ca/S ratio less than 1.5.

This phase was conducted with an Ohio #6 coal with an average sulfur content of 3.94% by weight. It produced about 3000 ppm of SO_2 at the scrubber inlet. Table 4 shows an average of all coal analyses for this phase. Coal sulfur varied from a minimum of 3.58% to a maximum of 4.30%.

The workscope for this phase greatly expanded when initial results showed that performance was well below the levels obtained in 1991. SO_2 removal was only 54% compared to 92% in 1991.

Sample analyses showed that the poor performance was likely caused by exceptionally fast evaporation in the scrubber as evidenced by the low moisture content of the baghouse ash and poor scrubber performance. At similar conditions, baghouse ash moisture was only 0.1% compared to 2.3% in 1991. These unexpected results shifted the focus from simply expanding the data base, to understanding the cause of the poor performance and finding a solution. Additional tests were conducted to resolve this issue and included changing atomizer operations (e.g. atomizing air, vent air, and atomizer type), reducing the limestone grind size, changing slaking techniques, and modifying baghouse operation. Finally, a deliquescent salt ($CaCl_2$) was added to the slurry to slow the rate of evaporation and increase the residual moisture in the baghouse ash.

TABLE 4   Average Coal Analyses for Phase 5

	As Fired	Dry
Gross Heating Value		
Btu/lb.	12595	13079
Btu/lb. (M&A Free)		14402
Ultimate Analysis, Wt. %		
Moisture	3.71	
Carbon	70.07	72.77
Hydrogen	4.90	5.09
Nitrogen	1.30	1.35
Sulfur	3.94	4.09
Ash	8.70	9.05
Oxygen (by difference)	7.33	7.65

Table 5 is a summary of important operating parameters and results for all LIDS testing, including four tests from Phase 6. Phase 6 results were included because these tests were conducted without ammonia injection and, therefore, can be considered LIDS tests. In all, 31 LIDS data sets were taken. The first column of the table gives a short description of each test. The other columns contain system temperatures and differential pressure, atomizer air and slurry flows, gas flow at the scrubber inlet, scrubber T_{as}, furnace limestone flow rate, furnace Ca/S ratio, SO_2 concentration at the inlet of each component adjusted to 3% oxygen, SO_2 removal across each component and across the LIDS system, and calcium utilization. It should be noted that some of the data for Test 7 may be misleading because this test was conducted at a T_{as} of 3°F and, consequently, the high moisture content of the material collected in the scrubber hopper lead to seemingly high values for slurry flow rate. Finally, a list of nomenclature is given below the table.

Table 5 Summary of Phase 5 and Phase 6 LIDS Tests

Phase 5 Test Description	Test	Temperature (°F) DSin	DSout	Twb	Differential Pressure (in.) HX	DS	BH	Atomizer Slurry (lb/hr)	% Solids	Atom Air (lb/hr)	Vent Air (lb/hr)	A:W (lb/lb)	DSin Gas (lb/hr)	Tas (°F)	LS (lb/hr)	SR SBS	SO₂ Data (ppm @ 3% O₂, dry) SBS	DSin	DSout	BHout	SO₂ Removal (%) SBS	DS	BH	Total	Ca Util.
Baseline	1	286	261	124	9.7	1.6	4.3	0	0.0	0	0	#N/A	4992	136.9	0.0	0.00	3290	3321	3412	3869	-0.9	-2.7	-13.4	-16.5	#N/A
Parametric	2	302	146	124	13.4	1.1	4.3	278	34.0	39	56	0.14	3970	21.9	71.3	1.82	3345	2391	1680	1532	28.5	29.7	8.8	54.2	29.8
Baseline	3	297	272	124	10.9	1.3	3.5	0	0.0	0	0	#N/A	5007	147.9	0.0	0.00	3207	3207	3422	3382	0.0	-6.7	1.2	-5.5	#N/A
Limestone Only	4	301	276	124	13.0	1.3	2.8	0	0.0	0	0	#N/A	4908	152.1	95.1	1.97	3224	2321	2448	2426	28.0	-5.5	0.9	24.8	12.6
Parametric	5	304	134	125	11.7	1.2	4.1	361	42.6	46	88	0.13	5129	9.4	90.0	1.96	2913	2293	1379	929	21.3	39.9	32.6	68.1	34.7
Low A/W Test,	6	297	133	124	11.8	1.3	3.9	353	37.8	27	90	0.08	5044	9.8	94.5	1.92	3192	2378	1310	823	25.5	44.9	37.2	74.2	38.6
Low A/W Test	7	299	128	125	13.3	1.5	3.9	475	37.8	25	93	0.05	4967	2.9	94.5	1.89	3212	2273	853	425	29.2	62.5	50.2	86.8	45.9
Aged Slurry	8	303	133	122	11.7	1.4	4.5	307	36.0	102	105	0.33	4962	10.9	90.0	1.92	3192	2311	1070	425	27.6	53.7	60.3	86.7	45.1
Aged Slurry	9	300	134	123	12.3	1.4	4.5	313	33.2	41	84	0.13	4998	10.9	90.0	1.91	3192	2216	890	442	30.6	59.8	50.4	86.2	45.1
Hot Slake	10	296	129	124	14.5	1.5	4.4	329	33.8	98	114	0.30	5126	5.1	90.0	1.80	3191	2285	1516	1117	28.4	33.7	26.3	65.0	36.0
Hot Slake	11	300	125	124	14.3	1.6	4.0	376	35.5	115	139	0.31	5122	1.1	90.0	1.83	3150	2363	1206	817	25.0	49.0	32.3	74.1	40.4
Parametric	12	298	136	126	13.9	1.7	3.7	328	39.7	105	129	0.32	5136	10.8	90.0	1.85	3109	2486	1821	1545	20.0	26.8	15.2	50.3	27.2
Parametric	13	300	130	125	13.8	1.5	4.5	338	40.1	111	133	0.33	5067	5.3	90.0	1.86	3130	2455	1573	1209	21.6	35.9	23.2	61.4	33.0
No DID	14	301	135	124	13.4	1.5	3.4	305	34.6	101	122	0.33	4956	11.9	90.0	1.93	3177	2302	1539	1174	27.5	33.1	23.7	63.0	32.7
No DID	15	300	133	125	11.5	1.4	5.8	338	34.6	70	118	0.21	4924	9.0	90.0	1.93	3177	2365	1268	829	25.6	46.4	34.6	73.9	38.3
Baseline	D1	304	266	#N/A	12.2	1.4	3.0	0	0.0	108	106	#N/A	4780	#N/A	0.0	0.00	2718	2852	2907	2961	-4.9	-1.9	-1.9	-8.9	#N/A
Limestone Only	D2	312	273	#N/A	16.7	1.5	4.0	0	#N/A	108	106	#N/A	4676	#N/A	81.0	2.23	2719	1919	1899	1982	29.4	1.0	-4.4	27.1	12.1
Fine Limestone	D3	300	133	122	18.3	1.4	3.9	373	42.0	106	90	0.28	4772	11.5	81.0	2.13	2717	1891	1050	752	30.4	44.5	28.4	72.3	33.9
BH Cycle	D4	300	130	122	19.4	1.5	2.6	358	42.0	109	89	0.31	4772	8.4	81.0	2.13	2717	1816	1081	771	33.2	40.5	28.6	71.6	33.6
Vent Air	D5	308	130	122	19.9	1.5	3.1	360	42.0	107	0	0.30	4772	7.7	81.0	2.13	2717	1890	1068	736	30.4	43.5	31.1	72.9	34.2
0.5% CaCl₂	D6	307	145	124	18.7	1.5	3.4	301	36.8	93	78	0.31	4791	20.8	81.0	2.08	2797	2087	1510	1474	25.4	27.6	2.4	47.3	22.7
1% CaCl₂	D7	299	130	121	18.2	1.4	4.9	362	37.9	95	83	0.26	4802	9.1	85.0	2.09	2873	1998	529	107	30.5	73.5	79.8	96.3	46.1
1% CaCl₂, Parametric	D8	301	127	121	19.4	1.4	4.5	372	38.5	95	84	0.26	4840	6.8	85.0	2.06	2939	1914	254	30	34.9	86.7	88.3	99.0	48.0
Baseline	D9	295	128	121	17.7	1.5	4.7	368	38.5	98	84	0.27	4875	7.3	0.0	0.00	2940	2970	1200	790	-1.0	59.6	34.2	73.1	#N/A
1% CaCl₂, Parametric	D10	299	129	120	18.1	1.4	5.0	326	33.2	101	84	0.31	4749	8.5	62.6	1.57	2883	2239	1055	21	22.3	52.9	98.0	99.3	63.1
0% CaCl₂, Parametric	D11	301	130	121	17.5	1.5	5.1	364	37.5	96	84	0.26	4900	8.2	62.6	1.52	2915	2310	1324	1014	20.8	42.7	23.4	65.2	42.8
CaO Scavenging	D12	301	130	120	19.7	1.4	3.1	340	37.5	95	83	0.28	4785	10.2	0.0	0.00	2954	2945	2182	1735	0.3	25.9	20.5	41.3	#N/A

Note: In Test 7, some of the data is misleading due to the high moisture content in the dry scrubber bottom ash.

Phase 6 Test Description	Test	DSin	DSout	Twb	HX	DS	BH	Slurry	% Solids	Atom Air	Vent Air	A:W	DSin Gas	Tas	LS	SR	SBS	DSin	DSout	BHout	SBS	DS	BH	Total	Ca Util.
Parametric, no NH₃	5	299	140	122	19.7	1.6	6.9	354	50.2	109	98	0.31	5045	17.9	61.8	1.18	3420	2262	1321	1157	33.9	41.6	12.4	66.2	56.1
Parametric, no NH₃	8	299	130	119	18.2	1.5	7.2	412	43.5	140	94	0.34	4991	10.6	73.4	1.44	3406	2361	646	16	30.7	72.7	97.5	99.5	69.0
Parametric, no NH₃	11	302	130	120	17.3	1.5	7.0	358	40.9	141	93	0.39	5130	10.1	48.9	1.00	3166	2602	1081	412	17.8	58.5	61.9	87.0	86.7
Parametric, no NH₃	12	298	140	120	16.6	1.5	8.1	312	42.9	96	104	0.31	5022	20.3	48.9	1.02	3164	2359	1751	1181	25.4	25.8	32.6	62.7	61.7

Nomenclature

Twb = Wet Bulb Temperature
HX = Heat Exchanger
DS = Dry Scrubber
BH = Baghouse
A:W = Air to Liquid Ratio of Atomizer

Tas = Approach to Saturation Temperature
LS = Limestone
SBS = Small Boiler Simulator - Test Furnace
SR = Calcium to Sulfur Molar Ratio
Util = Sorbent Utilization = SO₂ Removal/SR

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Baseline and Exploratory Tests

The first 5 runs were baseline tests and tests at normal LIDS conditions. At the completion of Test 5, it was obvious that performance goals could not be met, with only 54% SO_2 removal compared to 92% in 1991. Therefore, several exploratory tests were conducted in an attempt to improve performance. Tests 6 and 7 were run at very low air-to-slurry ratios to increase drop size and slow evaporation, but resulted in only slight improvement. Tests 8 and 9 were run with aged slurry and with different atomizers designed to operate at an A/W=0.3 lb/lb and 0.1 lb/lb. Results showed no difference due to atomizer type, but there was about a 12% improvement from Test 6 indicating that slaking technique may have had an effect on performance. Tests 10 and 11 used hot water for slaking. This improved hydration efficiency, but did not improve overall performance. Tests 12 and 13 were normal LIDS tests to compare with Test 5, but showed no improvement due to atomizer type. Tests 14 and 15 were run without the DID and with slurry aged for 48 hours. Again, no significant improvement was observed.

At this stage, it became obvious that changes in atomizer operating conditions and slaking technique were not sufficient to affect the needed improvement. Therefore, other avenues, involving reduced limestone grind size and the addition of a deliquescent salt ($CaCl_2$) to the slurry, were investigated. Several short tests were also done at this time, such as studying the effect of atomizer vent air, baghouse cleaning cycle, and water scavenging by CaO.

A scanning electron microscope revealed that particles in ash samples from the 1991 tests were much smaller than from Phase 5 samples. Subsequent particle size testing showed that although identical limestone was ordered from the same supplier, Phase 5 limestone had a larger mass mean diameter of 19.6 μ compared to 13.5 μ in 1991, which corresponds to a 45% increase in particle size and a 25% decrease in surface area. Therefore, a finer limestone was purchased for the remaining tests. The fine limestone had a mean diameter of 6.3 μ, which corresponds to a 53% decrease in particle size and a 52% increase in surface area from that used in 1991. However, furnace removal did not improve significantly (see Figure 8). Test 15 (coarse limestone) achieved 25.6% furnace SO_2 removal at a Ca/S ratio of 1.93 compared to 30.4% at 2.14 for Test D3 (fine limestone). Overall performance also did not improve with 74% SO_2 removal for Test 15 and 72% for Test D3.

Tests D4 and D5 were conducted to study the effect of the baghouse cleaning cycle and the quantity of atomizer vent air. Vent air affects the process by changing the entrainment characteristics of the atomizer. Flue gas entrainment decreases with increased vent air, and may affect flue gas mixing enough to decrease SO_2 removal. In Test D5, the vent air flow was terminated, but had no significant effect on performance. In Test D4, the baghouse cleaning cycle was decreased to 15 minutes to simulate the rate in 1991, but performance did not improve. Although these tests did not produce the required removal, the results were in good agreement with Test D3 demonstrating data reproducibility.

Effect of CaCl₂ on System Performance

Finally, the addition of a deliquescent salt ($CaCl_2$) was investigated. It was thought that the lack of chlorides in the coal was affecting scrubber performance and moisture levels in the baghouse ash by accelerating the evaporation rate. Chloride concentrations were measured in ash samples from 1991 LIDS tests, from Phase 4 and from Phase 5. The 1991 and Phase 4 samples had chloride levels averaging 0.14% Cl by weight in the dry ash, while Phase 5 samples had only 0.06%. Therefore,

$CaCl_2$ was added to the slaking tanks to increase the chloride concentration. With 1% by weight $CaCl_2$ addition, the concentration increased to 0.34%. Although this is greater than the concentration measured in 1991, it is well below the level where ash handling problems have been reported to occur[5], and no problems were observed.

Figure 7 shows SO_2 removal across the LIDS system for tests at a Ca/S of 2.0 with and without chloride addition. At a T_{as} of 10°F, tests without chloride addition averaged only 68%, while tests with 1% $CaCl_2$ addition achieved over 97% SO_2 removal. Although the addition of chlorides restored performance, it is believed that other unidentified combinations of factors, such as flyash chemistry or changes in the flow fields, also contributed to high evaporation rate. All remaining tests in Phase 5, and all tests in Phase 6, were conducted with 1% chloride addition. The effect of the chloride ion has been extensively researched in the industry[6,7,8].
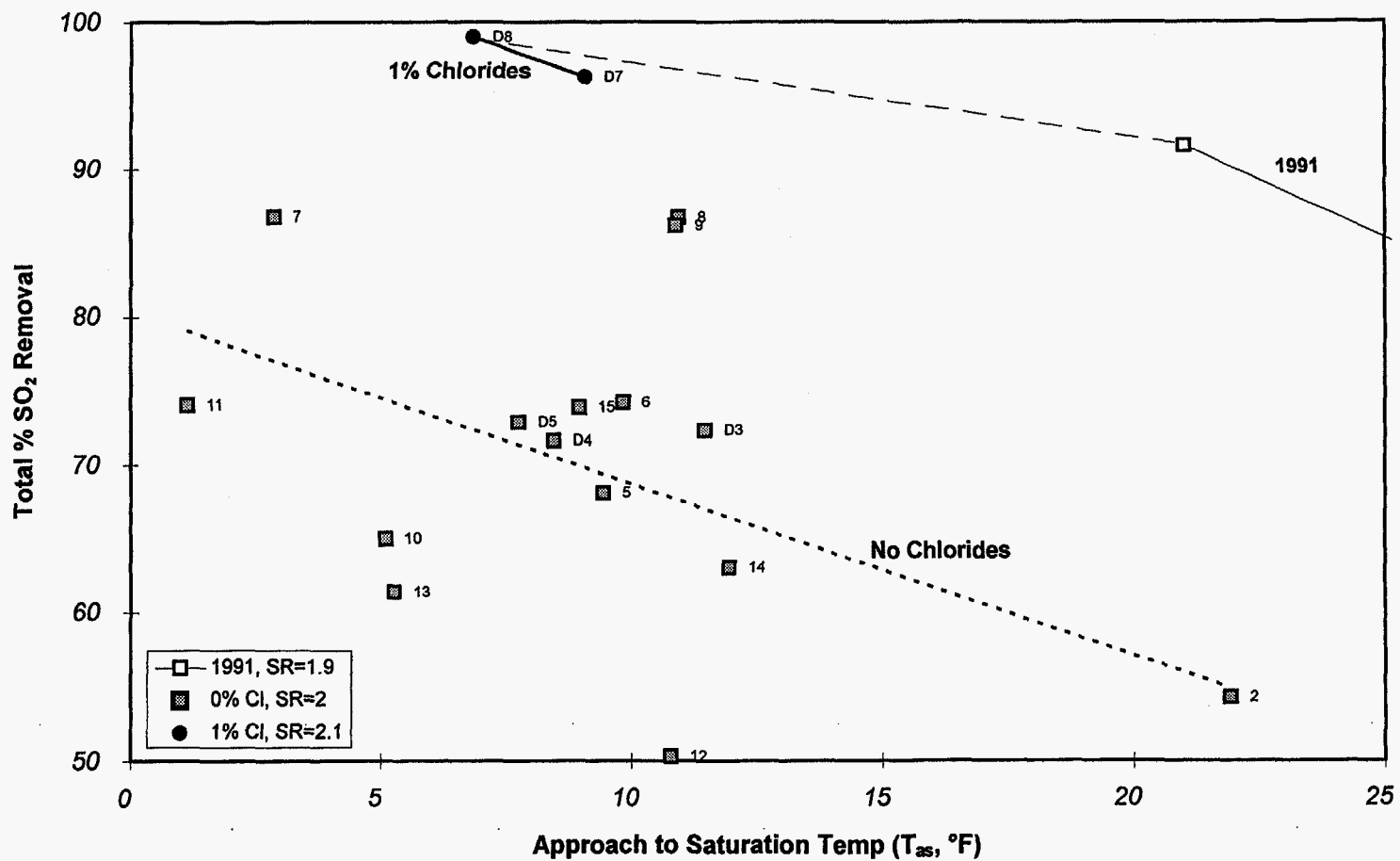
Furnace SO_2 Removal

Figure 8 shows furnace SO_2 removal as a function of furnace Ca/S ratio for Phase 5 and for select data from 1991. All data fall within a band typical of furnace injection processes. Chemical analyses showed that calcination was similar for all limestone grind sizes. It is interesting to note, however, that the residual moisture in the baghouse ash increased from less than 0.5% by weight with the coarser grind sizes, to 0.92% by weight for the fine grind (see Figure 12). This indicates that particle size affects evaporation, but was not solely responsible for the low moisture levels compared to 1991 tests. The remaining figures in this section represent only those tests with chloride addition. Several data points from the 1991 tests and from Phase 6 are also included.

Tests D7, D8, and D10 were run with 1% $CaCl_2$ addition and resulted in SO_2 removal above 97%. Test D9 was a baseline test with slurry injection, but without limestone injection, to establish SO_2 concentrations and calculate required limestone rates. For Test D10, the furnace Ca/S ratio was reduced from 2.1 to 1.6. Consequently, the dry scrubber removal was lower, but this was compensated for by very high removal in the baghouse (98%). This demonstrates the advantage of having a three-stage removal system in that unused sorbent from the furnace and scrubber is still available to react in the baghouse. Test D11 was run at the same conditions as D10 but without $CaCl_2$ addition to confirm the effect of chlorides. Total SO_2 removal decreased from 99% to 65%.

Dry Scrubber, Baghouse, and Overall SO_2 Removal

Figures 9-11 show SO_2 removal, adjusted to 3% oxygen, across the scrubber, baghouse, and total LIDS system. Data are grouped by nominal furnace Ca/S ratios of 1.1, 1.5, and 2.0. The dashed line is an extrapolation from 1991 results at a nominal Ca/S ratio of 2.0, and exemplifies the success of this phase in that LIDS was extended to higher SO_2 removal at lower T_{as} than was possible in 1991. The dotted line represents an average of all tests without chloride addition (Tests 1-D6). Tests from Phase 6, designated by "P6", include three points at a nominal Ca/S ratio of 1.1 and one point at a Ca/S of 1.44. The best performance achieved by the LIDS system was over 99% SO_2 removal at a Ca/S ratio of 1.44 and a T_{as} of 10.6°F. This exceeds the goal of 95% removal at a Ca/S ratio of 1.5.

FIGURE 7   Effect of CaCl₂ on Total LIDS Performance
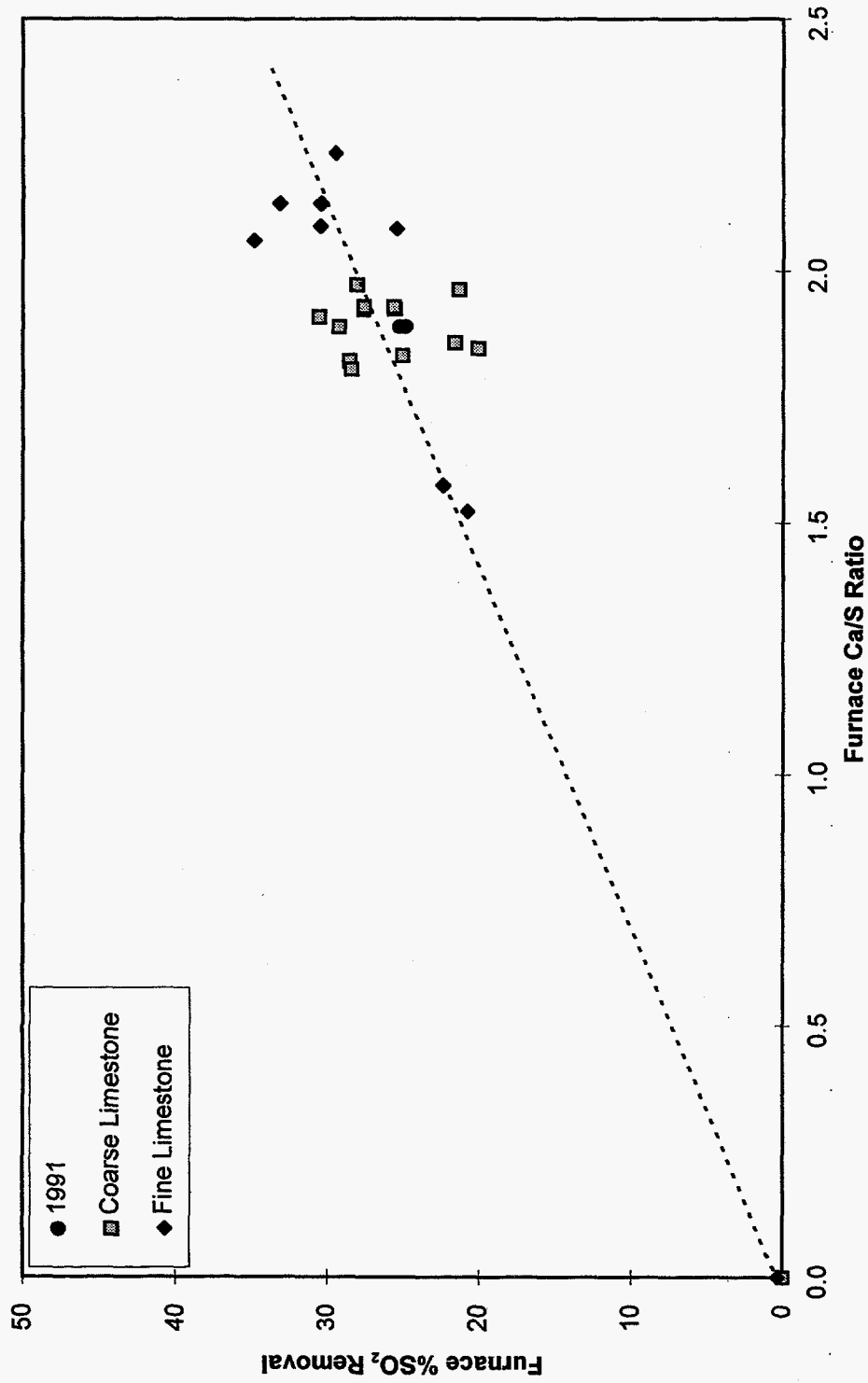
FIGURE 8   Phase 5 - Furnace SO₂ Removal
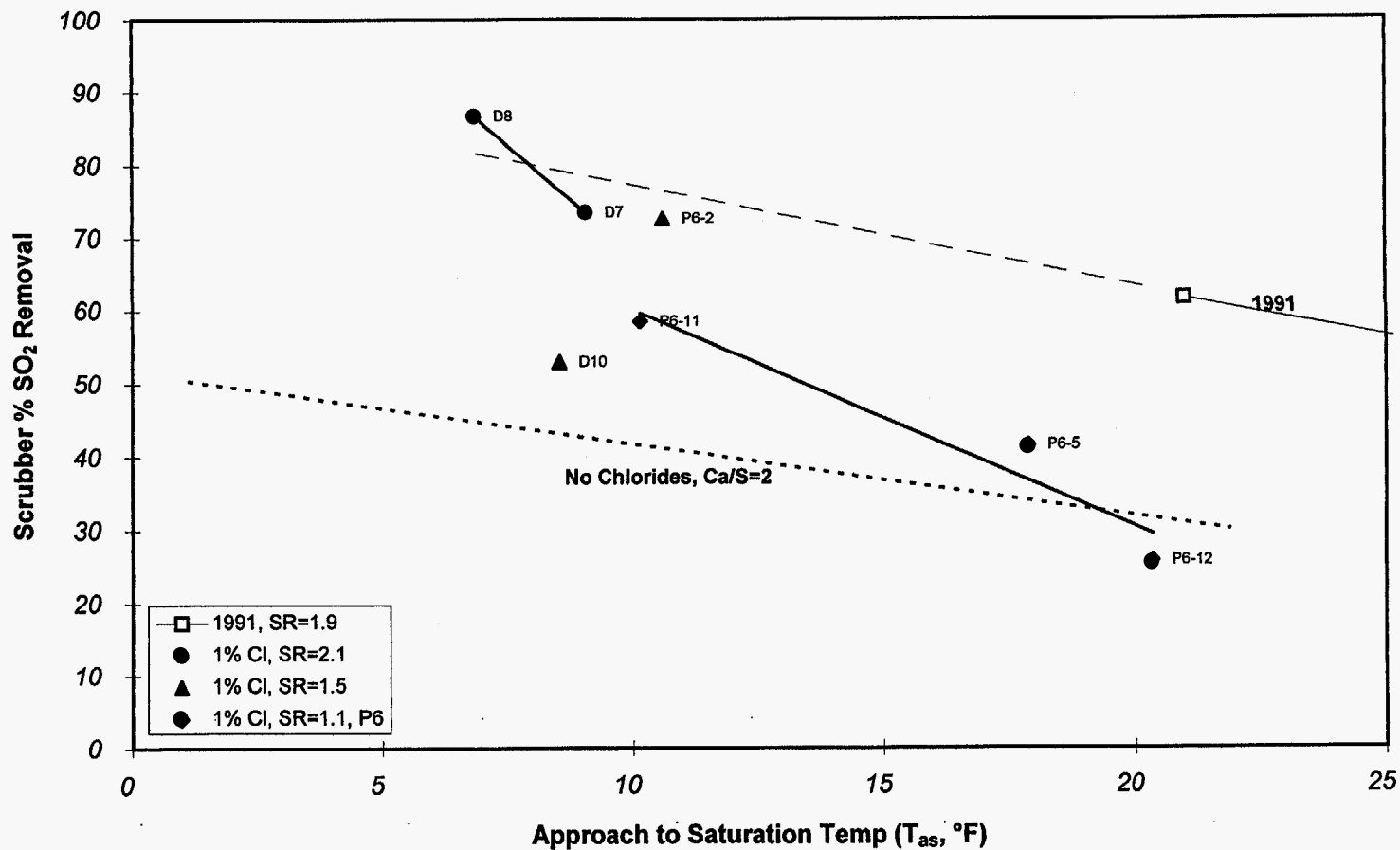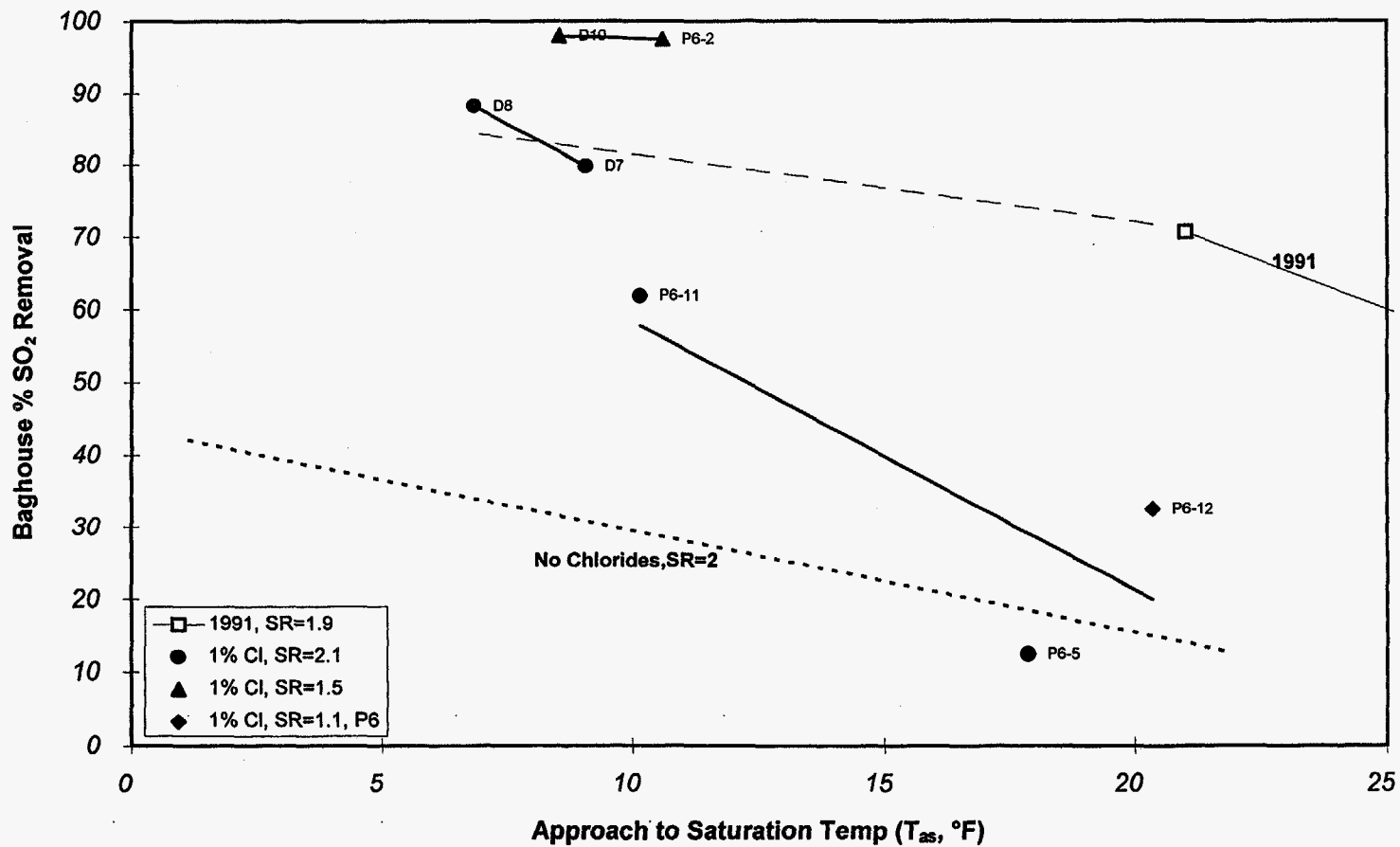
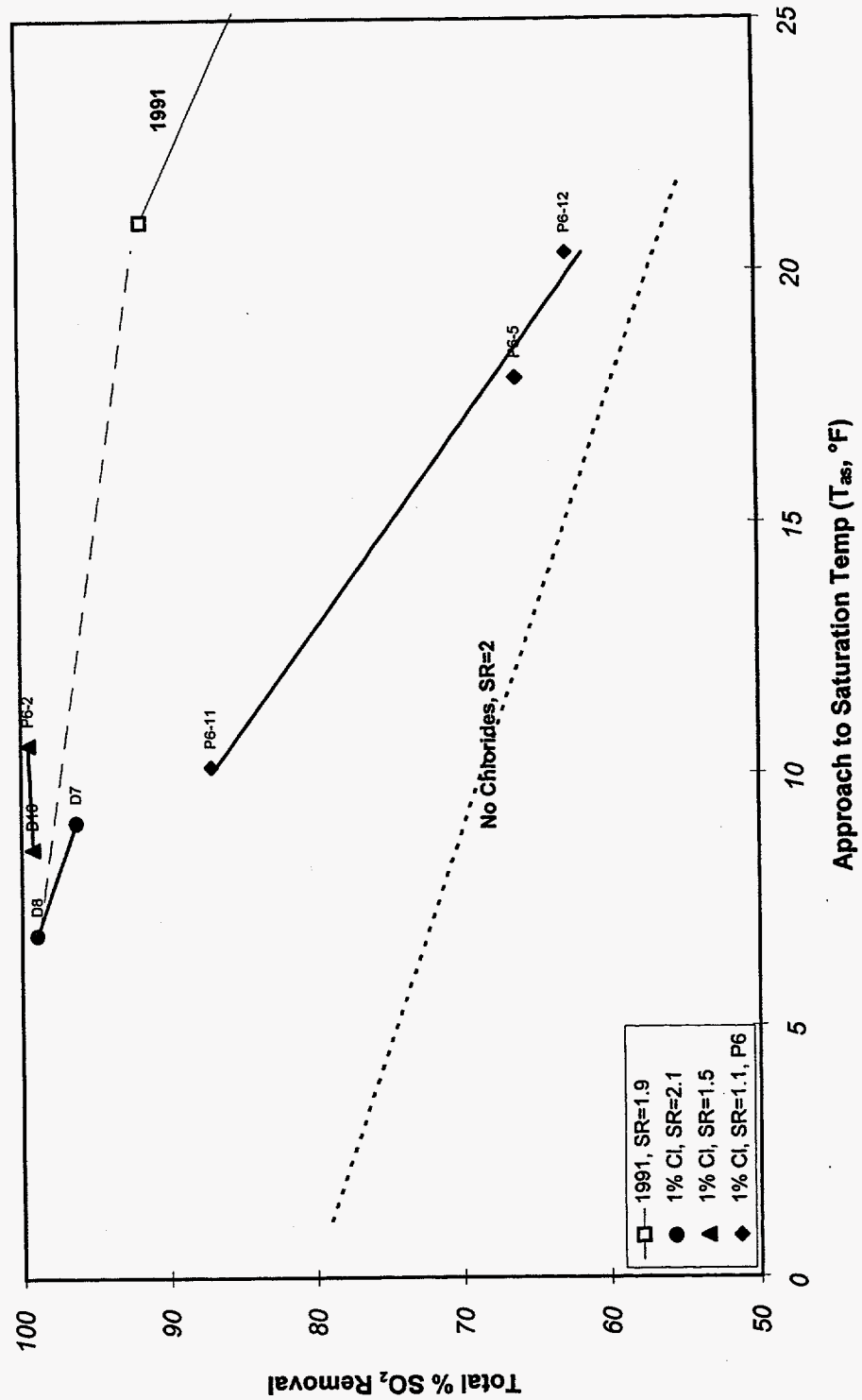## FIGURE 9   Phase 5 - LIDS Dry Scrubber SO$_2$ Removal

FIGURE 10   Phase 5 - LIDS Baghouse SO₂ Removal

FIGURE 11   Phase 5 - LIDS Total SO₂ Removal

An important discovery of this phase was the effect of residual baghouse ash moisture on scrubber and baghouse SO_2 removal. Figure 12 shows the dramatic relationship between ash moisture and baghouse SO_2 removal. Tests conducted without chloride addition, and with the coarser limestone, had moisture levels below 0.5% and SO_2 removal below 40%. With the addition of $CaCl_2$, the residual moisture increased to 1.6% and SO_2 removal increased to above 80%. This effect is not seen in dry scrubber systems that typically have higher ash moisture and show a direct relationship between T_{as} and ash moisture. The difference may be due to water scavenging by CaO from the furnace, and by the higher $Ca(OH)_2$ concentrations in dry scrubbing slurry. In 1991, the baghouse removal was 70% at a moisture level of 2.3%, but this was at a T_{as} of 20°F compared to 10°F for the recent tests. The exact cause of the vastly different performance and ash moisture between tests in 1991 and 1994 was not discovered.

The final test, D12, studied the effect of water scavenging by CaO. Mass balances showed that the ratio of CaO to water in the scrubber was slightly greater during Phase 5 than in 1991 because of higher coal sulfur. CaO reacts readily with water to form $Ca(OH)_2$, and it was thought that this may contribute to the high evaporation rate and low-ash moisture. To test this theory, the limestone feed to the furnace was discontinued to eliminate the source of new CaO. Slurry flow to the dry scrubber continued at the same rate. Baghouse ash samples were analyzed for moisture every half-hour. Before the test began, ash moisture was 0.18%. After five hours, the moisture had risen to about 2.1% and remained steady. Assuming that the evaporation rate is not affected significantly by the decrease in particulate loading, calculations show that the moisture content, due to decreased particulate loading, should less than double. Since the moisture increased by a factor of 12, it is likely that water is being scavenged by CaO. Based on the amount of limestone injected and assuming 80% is calcined, the difference in moisture indicates that about 20% of the CaO hydrates in the scrubber and baghouse. This confirmed that CaO scavenging occurs and may significantly affect baghouse ash moisture and SO_2 removal.

Post-test inspections revealed no significant deposition on the atomizer, scrubber walls, flues, or in the baghouse. The ash from the baghouse remained dry and free-flowing for all tests. This demonstrates the potential of B&W's dry scrubber advancements in atomizer design and deposition control to enable future dry scrubbers to operate cleanly at very low T_{as}.

2.2.5   Phase 6 - A+LIDS

A+LIDS, Ammonia and Limestone Injection Dry Scrubbing, (US Patent 5,176,088) is a new dry-scrubber-based technology developed by B&W. A+LIDS combines furnace-ammonia injection, also known as selective non-catalytic reduction (SNCR), with LIDS to expand dry scrubbing into the area of NO_x removal, and increase system SO_2 removal and sorbent utilization. SNCR and LIDS have

FIGURE 12 Effect of Baghouse Ash Moisture on SO₂ Removal
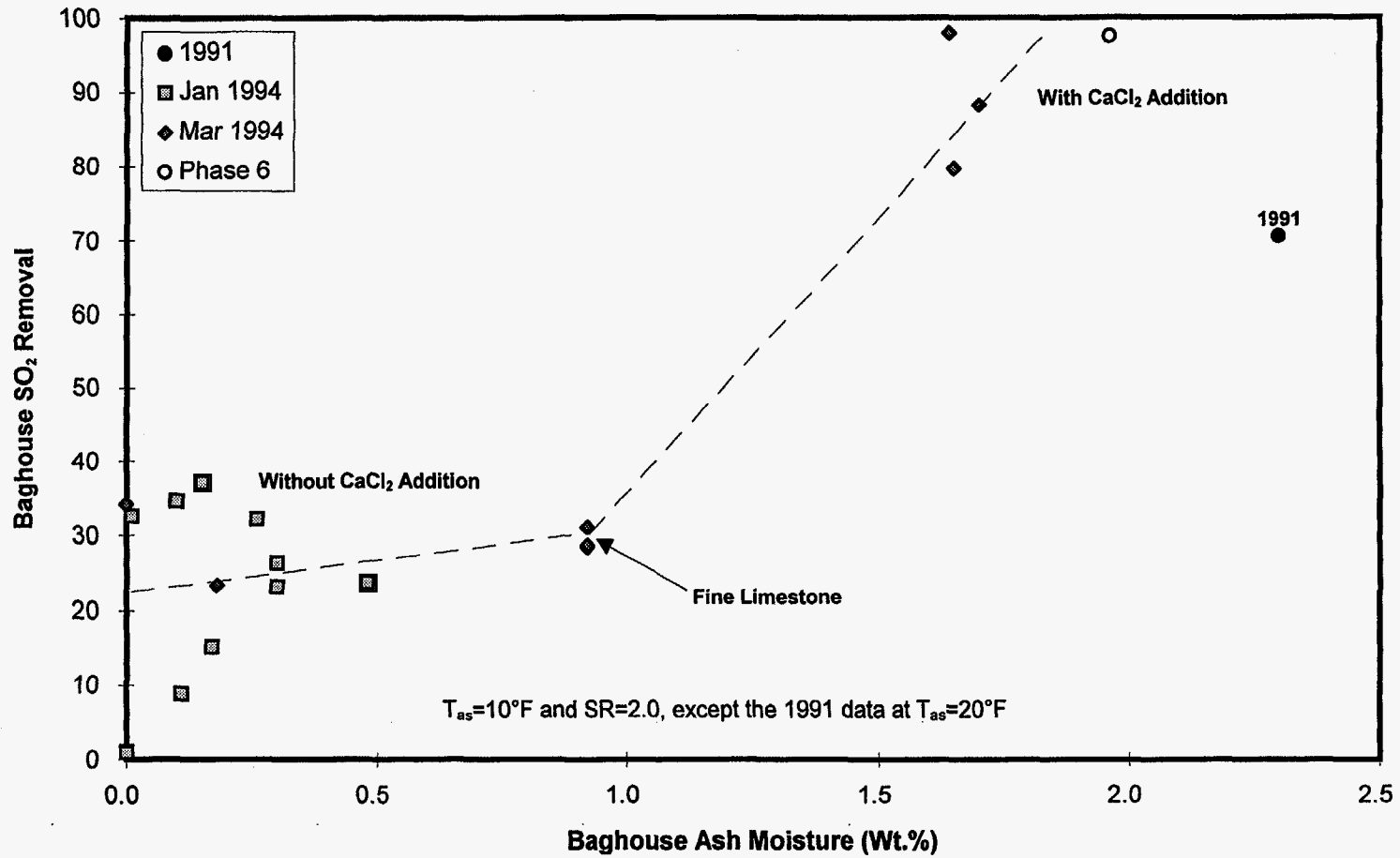
been successfully demonstrated separately but have never been combined. The advantages of combining these technologies are not readily apparent and go beyond what is possible with the individual technologies.

1. 95+% SO_2 removal.
2. 60+% NO_x removal if combined with low NO_x burners.
3. Low cost sorbents (ammonia and limestone).
4. No bisulfate fouling of the air heater, which is a possible consequence of SNCR.
5. No SO_3 condensation in the air heater or flues.
6. Control of ammonia slip.
7. The ability to maintain high SO_2 removal at higher scrubber approach temperatures.
8. High sorbent utilization.
9. Increased heat cycle efficiency due to a lower acid dew point.

A schematic of A+LIDS is shown in Figure 13. A+LIDS uses both limestone and ammonia injection, so the reaction paths of these reagents will be discussed separately. The calcium path is the same as for LIDS and begins with limestone injection into the upper furnace. Limestone calcines to lime and reacts with a portion of the SO_2 in the flue gas. Lime also reacts with SO_3, which prevents ammonium bisulfate formation in the air heater, a consequence of ammonia injection, and lowers the acid dew point. Unreacted lime and other particulate pass through the system to the baghouse where a portion is removed and slaked to make slurry for the dry scrubber. The lime in the slurry provides SO_2 removal in the dry scrubber and baghouse similar to conventional dry scrubbing and LIDS.

Furnace limestone injection is closely followed by the addition of ammonia to control NO_x. More than the stoichiometric requirement of ammonia can be added to increase NO_x removal and inhibit ammonium bisulfate formation by kinetically favoring ammonium sulfate formation. Any unreacted ammonia passes through the system to the dry scrubber where it reacts quantitatively with SO_2 in the cool, humid environment. The reaction produces extremely high ammonia utilization as long as some SO_2 remains, which eliminates ammonia slip.

In an alkaline environment, calcium displaces the ammonia in ammonia salts releasing ammonia gas. Therefore, ammonia can be recovered from the waste ash by mixing the ash with a small quantity of water, as would normally be done to control dusting in an ash-disposal system. Recovered ammonia could be returned to the scrubber or furnace to improve sorbent utilization.

The important parameters in this process are the limestone and ammonia injection temperatures, the rate of ammonia and limestone injection, and the scrubber and baghouse T_{as}. The optimum temperature range for limestone injection is the same as for LIDS, 2100-2300°F. As in dry scrubbing, scrubber and baghouse SO_2 removal efficiency increase as T_{as} decreases, as does potential problems with deposition and ash handling. The optimum temperature for NO_x reduction is about 1800°F. Injection at higher temperatures causes ammonia to decompose to NO_x, which is undesirable since NO_x reduction is the purpose of SNCR. Injection at lower temperatures increases ammonia slip due to incomplete reactions. Ammonia slip is undesirable because it is a regulated pollutant and can lead to ammonium bisulfate (NH_4HSO_4) formation. Ammonium bisulfate is very corrosive and condenses at temperatures below 350°F, as found in most air heaters. The formation of ammonium

A⁺ LIDS

FIGURE 13 A+LIDS Process Schematic

Stack

Baghouse

Dry scrubber

Heat exchanger

Boiler

NH₃

Coal

Water

Waste disposal

NH₃ regeneration system

Reagent preparation system

Limestone receiving and preparation

NH₃
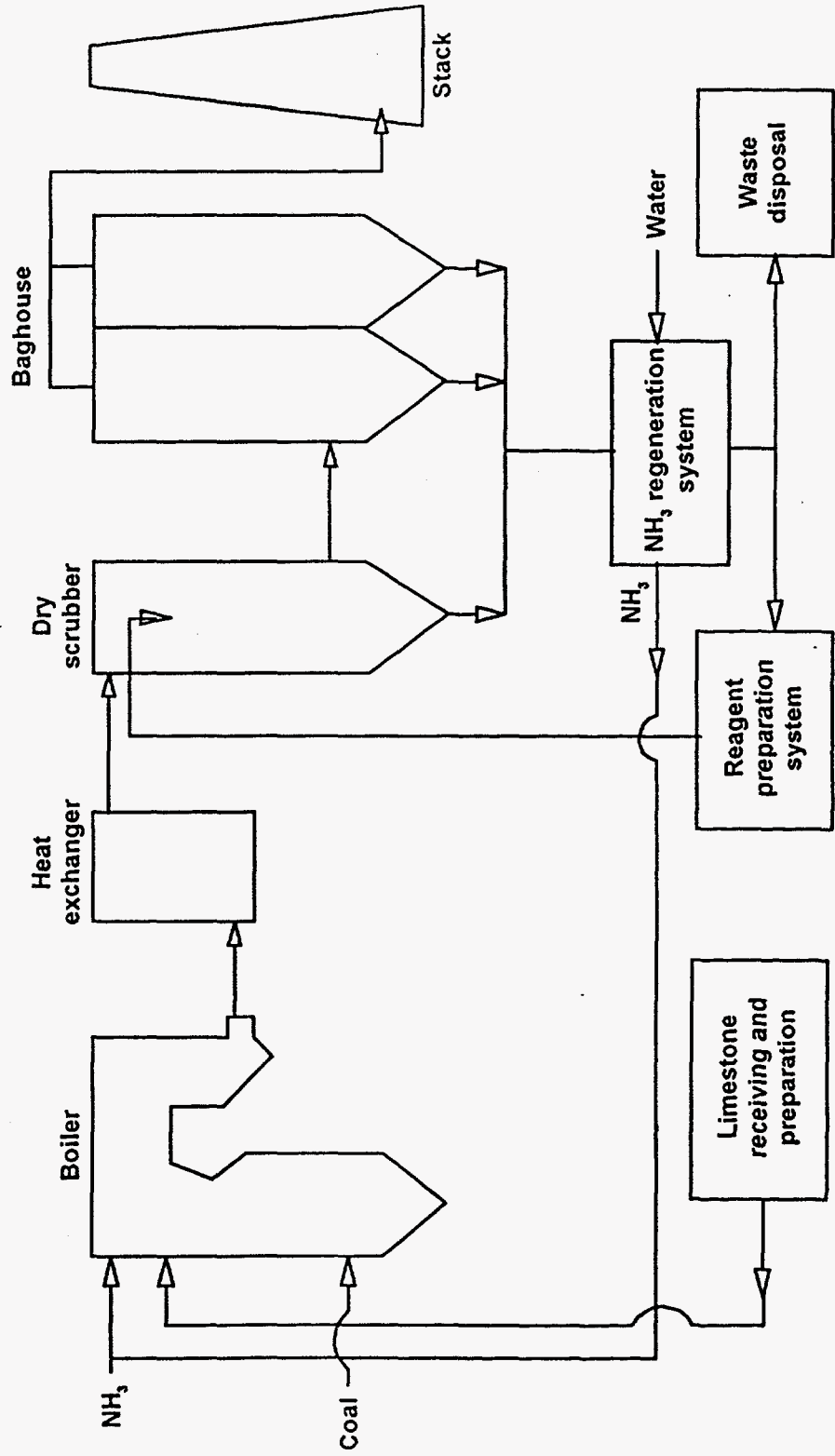
bisulfate can be controlled by reducing the SO_3 concentration through limestone injection, or by having a high excess of ammonia. A high excess of ammonia favors ammonium sulfate ($[NH_4]_2SO_4$) formation which does not lead to air heater fouling. The effect of ammonia addition on overall process efficiency was the major focus of this test phase.

The major chemical reactions for the A+LIDS technology are shown below. For greatest efficiency, the scrubber and baghouse calcium reactions must take place in the liquid phase. Once the slurry droplet has dried, the reactions no longer proceed.

In-Furnace Reactions - 1800-2000°F:
$$4NH_3 + 4NO + O_2 \rightarrow 4N_2 + 6H_2O$$
$$CaCO_3 + Heat \rightarrow CaO + CO_2$$
$$CaO + SO_2 + \frac{1}{2}O_2 \rightarrow CaSO_4$$
$$CaO + SO_3 \rightarrow CaSO_4$$
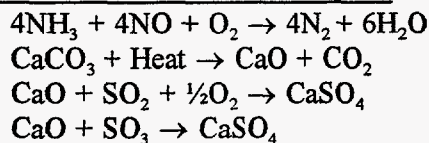
Air Heater Ammonia Reactions - <350°F:
$$NH_3 + SO_3 + H_2O \rightarrow NH_4HSO_4 \text{ (undesirable)}$$
$$2NH_3 + SO_3 + H_2O \rightarrow (NH_4)_2SO_4 \text{ (desirable)}$$
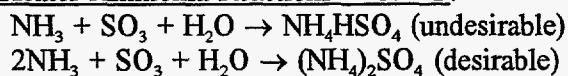
Dry Scrubber and Baghouse Ammonia Reactions - <300°F:
$$2NH_3 + H_2O + SO_2 \rightarrow (NH_4)_2SO_3$$
$$NH_3 + H_2O + SO_2 \rightarrow NH_4HSO_3$$

Ammonia Recovery:
$$Ca(OH)_2 + NH_4HSO_3 \rightarrow CaSO_3 + NH_3 + 2H_2O$$
$$Ca(OH)_2 + (NH_4)_2SO_3 \rightarrow CaSO_3 + 2NH_3 + 2H_2O$$
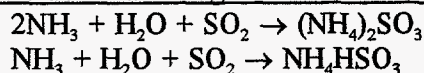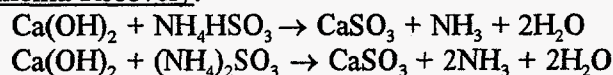
Slaking Reaction - Ambient:
$$CaO + H_2O \rightarrow Ca(OH)_2 + Heat$$

Scrubber and Baghouse Calcium Reactions:
$$Ca(OH)_2 \rightarrow Ca^{+2} + 2OH^-$$
$$SO_{2(aq)} + H_2O \rightarrow HSO_3^- + H^+$$
$$SO_{2(aq)} + OH^- \rightarrow HSO_3^-$$
$$HSO_3^- + OH^- \rightarrow SO_3^{-2} + H_2O$$
$$Ca^{+2} + SO_3^{-2} + \frac{1}{2}H_2O \rightarrow CaSO_3 \cdot \frac{1}{2}H_2O$$
$$\overline{Ca(OH)_2 + SO_2 \rightarrow CaSO_3 \cdot \frac{1}{2}H_2O + \frac{1}{2}H_2O}$$
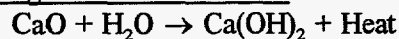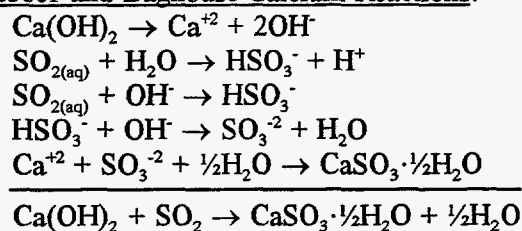
2.2.6 Phase 6 - Results

The main objective of this phase was to demonstrate the synergistic aspects of this state-of-the-art technology including high NO_x removal, high SO_2 removal, ammonia regeneration, and high calcium utilization. The goals for this phase were 60% NO_x removal, 95% SO_2 at Ca/S ratios less than 1.2, and greater than 90% ammonia recovery.

This phase was conducted with an Ohio #6 coal having an average sulfur content 3.99% as shown in Table 6. It produced about 3200 ppm of SO_2 at the scrubber inlet. The NO_x level at the burner outlet was maintained at about 630 ppm. Coal sulfur varied throughout the test, starting at 3.5%, peaking to 4.4%, and then returning to 3.9%. This variability made it difficult to maintain target Ca/S ratios.

TABLE 6   Average Coal Analyses for Phase 6

	As Fired	Dry
Gross Heating Value		
Btu/lb	12527	12912
Btu/lb (M&A Free)		14407
Ultimate Analysis, Wt.%		
Moisture	2.98	
Carbon	68.44	70.54
Hydrogen	5.02	5.17
Nitrogen	1.39	1.43
Sulfur	3.87	3.99
Ash	10.07	10.38
Oxygen (by difference)	8.22	8.47

Table 7 summarizes important operating parameters and performance results. The upper portion contains select flue gas temperatures, pressure drop across the major components, atomizer slurry and air flows, flue gas flow at the scrubber inlet, furnace limestone flow and Ca/S ratio, scrubber T_{as}, and the residual moisture in the baghouse waste. The lower portion contains SO_2 concentrations adjusted to 3% oxygen, SO_2 removal across each component, calcium utilization, furnace ammonia flow and NH_3/NO_x ratios, furnace NO_x removal, and ammonia slip measured at the baghouse outlet. No NO_x removal occurred across the scrubber or baghouse. Data are grouped in pairs corresponding to similar test conditions with and without ammonia injection. Thirteen tests were conducted including five baseline tests, four LIDS tests, and four A+LIDS tests. A list of nomenclature is also given.

Tests were conducted with and without ammonia injection in the furnace to study the synergistic effects of ammonia addition. Furnace Ca/S ratios ranged from 1.00 to 1.48, T_{as} ranged from 10°F to 20°F, and furnace NH_3/NO_x ratios ranged from 1.2 to 3.2. Ammonia flow was set by increasing the injection rate until an ammonia slip of less than 15 ppm could be maintained at the baghouse outlet. Because ammonia is absorbed in the dry scrubber, low levels of slip could be maintained while injecting excess ammonia in the furnace. Slip was controlled by simply reducing the amount of excess ammonia fed to the furnace. All tests were conducted with the addition of 1% by weight $CaCl_2$ to the slurry solids. The DID was installed throughout the phase but remained deposit free because of the high solids content of the slurry.

TABLE 7   Summary of Phase 6 - A+LIDS

Table 7 Summary of Phase 6 - A+LIDS

Test Description	Test	Temperature (°F)			Differential Pressure (in.)			Atomizer		Atom Air (lb/hr)	Vent Air (lb/hr)	A:W (lb/lb)	DSin Gas (lb/hr)	SBS LS (lb/hr)	SBS SR Ca/S	T_as (°F)	BH Ash %H₂O
		DSin	DSout	Twb	HX	DS	BH	Slurry (lb/hr)	% Solids								
Baseline	1	298	153	122	13.3	1.4	4.4	136	0.0	98	100	0.71	4746	0.0	0.00	31	0.00
Baseline	2	298	170	121	15.7	1.7	5.1	108	0.0	99	99	0.92	4821	0.0	0.00	49	0.00
Baseline	3	290	159	121	12.4	1.6	2.9	130	0.0	96	98	0.74	5067	0.0	0.00	38	0.00
LS Only	4	303	159	120	14.6	1.6	6.7	151	0.0	98	98	0.65	4908	61.8	1.37	39	0.23
Baseline	6	298	138	120	19.8	1.6	7.0	346	0.0	128	91	0.37	5006	0.0	0.00	18	0.00
Parametric, no NH₃	5	299	140	122	19.7	1.6	6.9	354	50.2	109	98	0.31	5045	61.8	1.18	18	0.53
Parametric, w/NH₃	7	307	140	120	20.3	1.6	5.5	376	50.2	116	91	0.31	5282	73.4	1.35	20	0.47
Parametric, no NH₃	8	299	130	119	18.2	1.5	7.2	412	43.5	140	94	0.34	4991	73.4	1.44	11	1.96
Parametric, w/NH₃	9	304	128	118	18.8	1.5	6.7	386	46.9	143	96	0.37	4973	73.4	1.48	10	2.60
Parametric, w/NH₃	10	303	130	120	17.0	1.6	7.9	412	45.9	139	93	0.34	5160	48.9	1.00	10	1.58
Parametric, no NH₃	11	302	130	120	17.3	1.5	7.0	358	40.9	141	93	0.39	5130	48.9	1.00	10	1.25
Parametric, no NH₃	12	298	140	120	16.6	1.5	8.1	312	42.9	96	104	0.31	5022	48.9	1.02	20	0.44
Parametric, w/NH₃	13	301	139	121	17.1	1.4	6.4	340	44.2	89	106	0.26	4944	48.9	1.08	17	0.06

Test Description	Test	SO₂ Data (ppm @ 3% O₂, dry)				SO₂ Removal (%)				%Ca Util.	Ammonia & NOₓ Data in the Furnace					NH₃ Slip (ppm)
		SBS	DSin	DSout	BHout	SBS	DS	BH	Total		NH₃ (lb/hr)	SR NH₃/NOₓ	SBS (ppm @ 3% O₂, dry)	DSin	% Rem.	
Baseline	1	3003	3133	3067	3026	-4.3	2.1	1.3	-0.8	#N/A	0.00	0.00	721	721	0.0	0
Baseline	2	2940	2721	2611	2654	7.5	4.0	-1.7	9.7	#N/A	0.00	0.00	614	614	0.0	0
Baseline	3	2948	2958	2510	2730	-0.3	15.1	-8.8	7.4	#N/A	0.00	0.00	658	658	0.0	0
LS Only	4	3108	1855	1674	1624	40.3	9.7	3.0	47.8	34.9	0.00	0.00	660	660	0.0	0
Baseline	6	3583	3458	2282	1924	3.5	34.0	15.7	46.3	#N/A	0.00	0.00	612	612	0.0	0
Parametric, no NH₃	5	3420	2262	1321	1157	33.9	41.6	12.4	66.2	56.1	0.00	0.00	619	619	0.0	0
Parametric, w/NH₃	7	3540	2479	1173	1199	30.0	52.7	-2.2	66.1	49.0	2.69	1.60	654	375	42.7	14
Parametric, no NH₃	8	3406	2361	646	16	30.7	72.7	97.5	99.5	69.0	0.00	0.00	573	573	0.0	5
Parametric, w/NH₃	9	3271	2498	867	6	23.7	65.3	99.3	99.8	67.5	2.16	1.52	561	403	28.1	8
Parametric, w/NH₃	10	3157	2755	589	31	12.7	78.6	94.8	99.0	99.3	1.70	1.15	569	360	36.7	14
Parametric, no NH₃	11	3166	2602	1081	412	17.8	58.5	61.9	87.0	86.7	0.00	0.00	628	628	0.0	0
Parametric, no NH₃	12	3164	2359	1751	1181	25.4	25.8	32.6	62.7	61.7	0.00	0.00	620	620	0.0	0
Parametric, w/NH₃	13	3048	2079	1046	863	31.8	49.7	17.4	71.7	66.4	4.98	3.21	626	273	56.4	7

Note: Negative removals indicate analyzer drift or differences in analyzer response time due to sampling location.

Nomenclature

Twb = Wet Bulb Temperature
HX = Heat Exchanger
DS = Dry Scrubber
BH = Baghouse
SBS = Small Boiler Simulator - Test Furnace

A:W = Air to Liquid Ratio of Atomizer
T_as = Approach to Saturation Temperature
LS = Limestone
SR = Ca(OH)₂/SO₂ or NH₃/NOₓ Molar Ratio
Util. = Calcium Utilization = Removal/SR
Rem. = Removal

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Furnace SO₂ Removal

Figure 14 shows furnace SO_2 removal as a function of furnace Ca/S ratio for Phase 6 and select data from the 1991 LIDS Demonstration. The figure shows that furnace SO_2 removal averaged about 5% higher than in 1991. The increase is likely due to improvements in the injection system that permitted greater sorbent coverage at the injection plane, and to the use of finer limestone (6 μ instead of 13 μ). Fine limestone was used because it exhibited better performance and moisture holding characteristics during Phase 5. At a Ca/S of 1.0, about 20% SO_2 removal was achieved. This shows that the major function of furnace injection is not SO_2 removal, but to produce a highly reactive waste suitable for dry scrubbing.
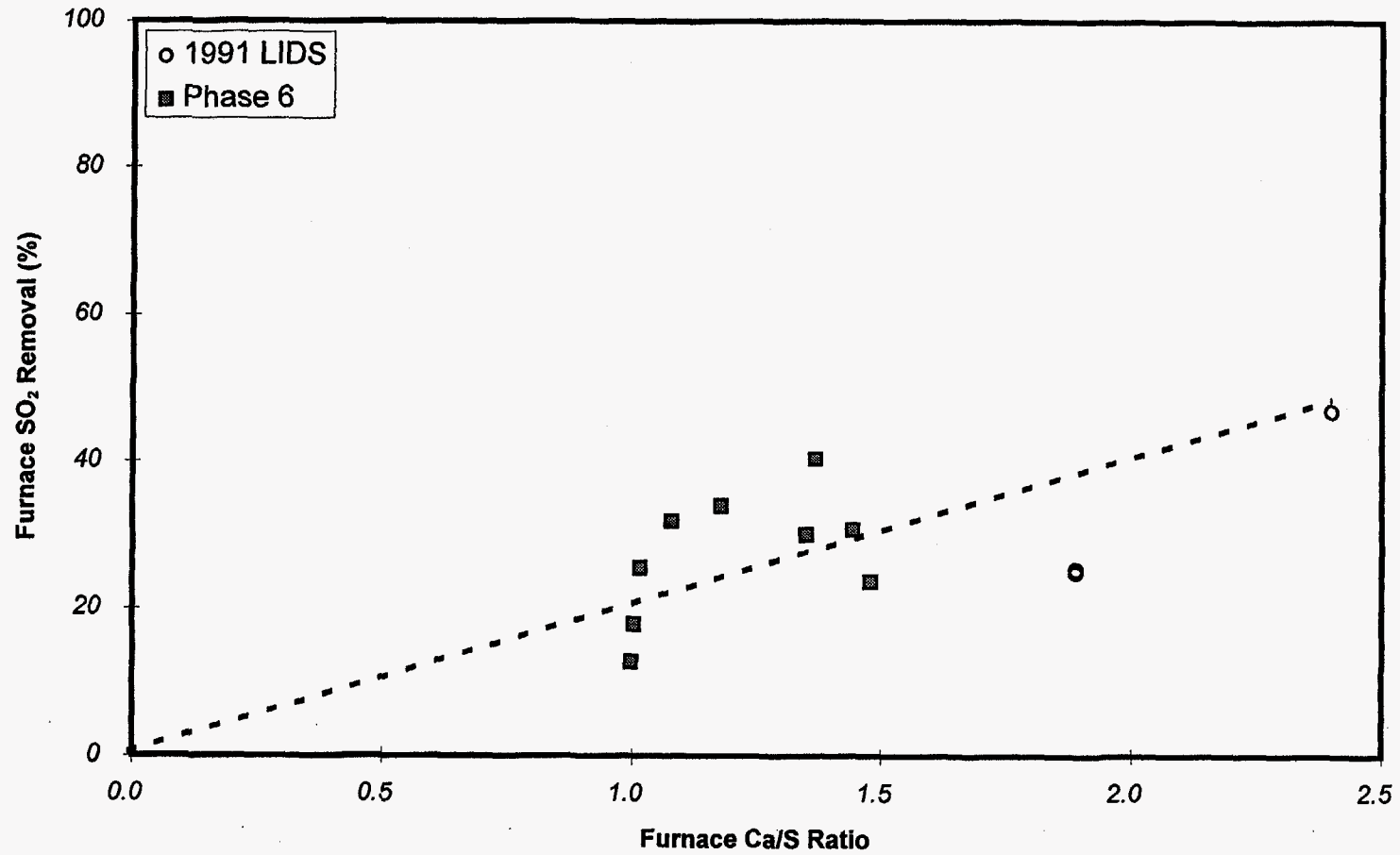
Dry Scrubber Performance

Figure 15 shows dry scrubber SO_2 removal as a function of T_{as}. Data are grouped by nominal furnace Ca/S ratios of 1.00, 1.22 and 1.44 and whether or not ammonia addition was employed. Data from 1991 at a Ca/S ratio of 1.9 are also included as a reference. At a T_{as} of 10°F and a Ca/S of 1.0, SO_2 removal across the scrubber increased from 59% to 79% when ammonia injection was used. The data point representing a T_{as} of 10°F and a Ca/S of 1.44 appears irregular, but no reason could be found for the low SO_2 removal. However, the total system SO_2 removal for the same test was not atypical because high removal occurred in the baghouse (Fig. 16). This demonstrates the advantage of multiple reaction sites, in that poor performance in one area is compensated in another.

Baghouse Performance

Figure 16 shows baghouse SO_2 removal as a function of dry scrubber T_{as}. Data are grouped by nominal furnace Ca/S ratios of 1.00, 1.22 and 1.44 and whether or not ammonia addition was employed. Data from 1991, at a Ca/S ratio of 1.9, are also included as a reference. At a T_{as} of 10°F and a Ca/S of 1.0, SO_2 removal across the baghouse increased from 62% to 95% when ammonia injection was used. However, this large effect was not evident in the other tests. In general, the figure shows that a baghouse is a very efficient method of removing SO_2 at low T_{as} (i.e., high residual moisture in the baghouse ash), regardless of whether or not ammonia is present. The data point representing a T_{as} of 20°F and a Ca/S of 1.0 appears irregular, but no reason could be found for the high SO_2 removal.

Figure 17 shows the effect of residual moisture in the baghouse ash on the SO_2 removal across the baghouse. As discovered during Phase 5, the figure shows that residual moisture has a significant impact on SO_2 removal for a LIDS-based system. When ash moisture was below 1.0%, SO_2 removal was below 40%. When ash moisture increased to above 1.5%, SO_2 removal increased above 90%. If low moisture becomes a problem on a commercial application, a method could be devised to automatically vary the concentration of a deliquescent material (such as calcium chloride) to maintain ash moisture at 1.5% and maximize the potential of the baghouse.

FIGURE 14   Phase 6 - Furnace SO$_2$ Removal

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FIGURE 15   Phase 6 - Dry Scrubber SO_2 Removal

Y-axis: Scrubber % SO_2 Removal

X-axis: Approach to Saturation Temp (T_{as}, °F)

Legend:
- 1991 LIDS
- SR=1.22 w/o NH3
- SR=1.44 w/NH3
- SR=1.00 w/o NH3
- SR=1.00 w/NH3

1991   SR=1.9
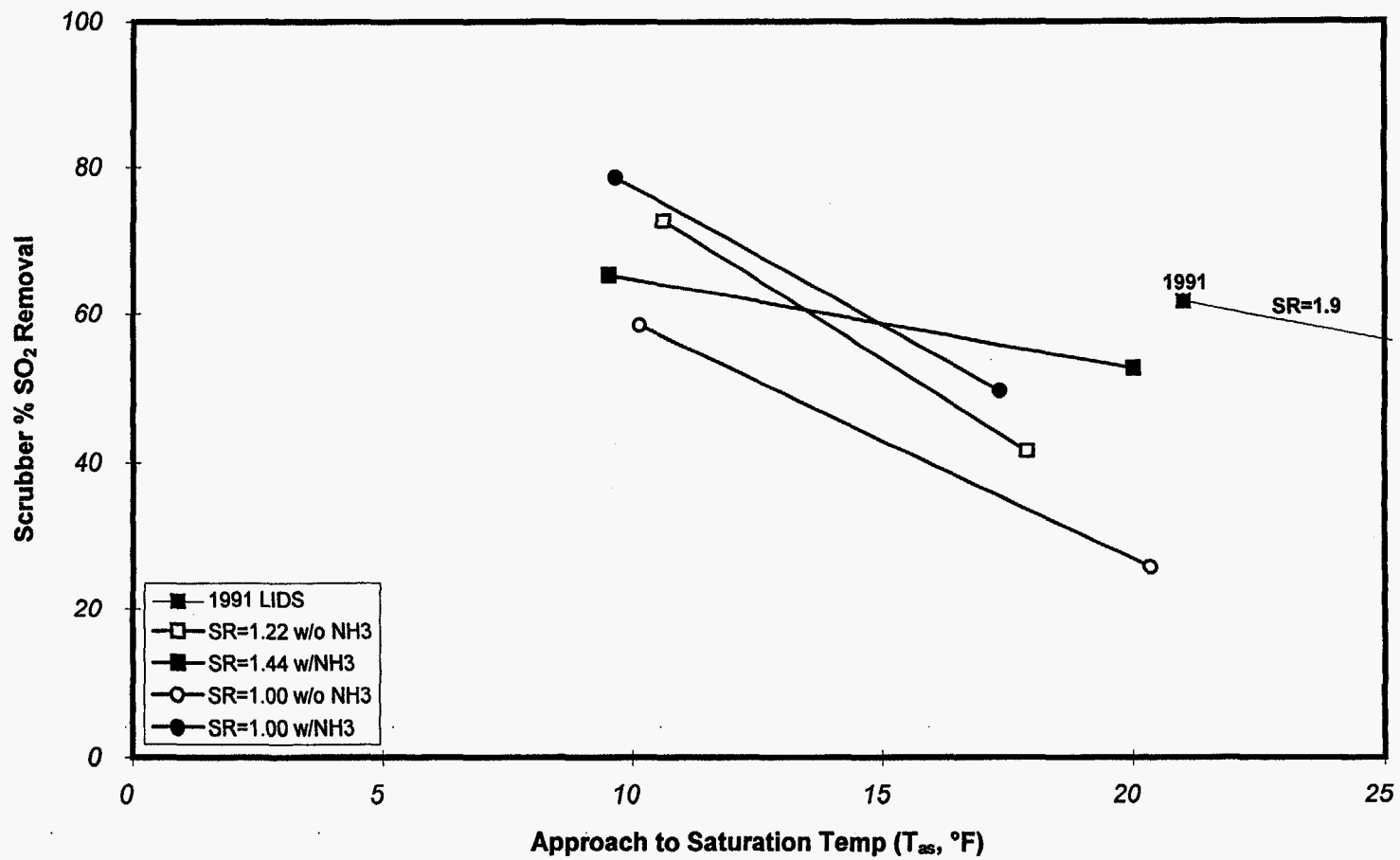
FIGURE 16 Phase 6 - Baghouse SO$_2$ Removal

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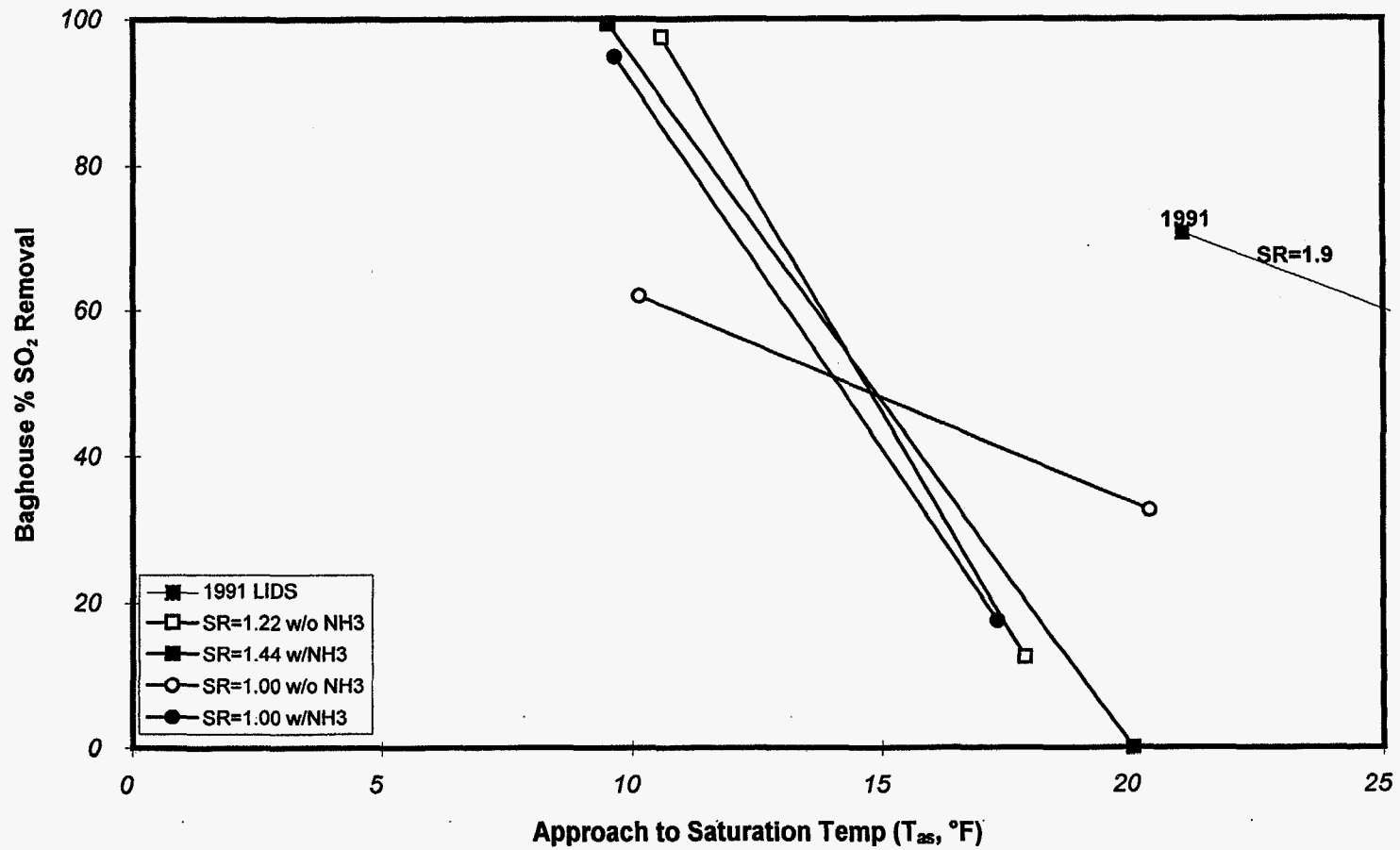
FIGURE 17 Effect of Baghouse Ash Moisture on SO₂ Removal
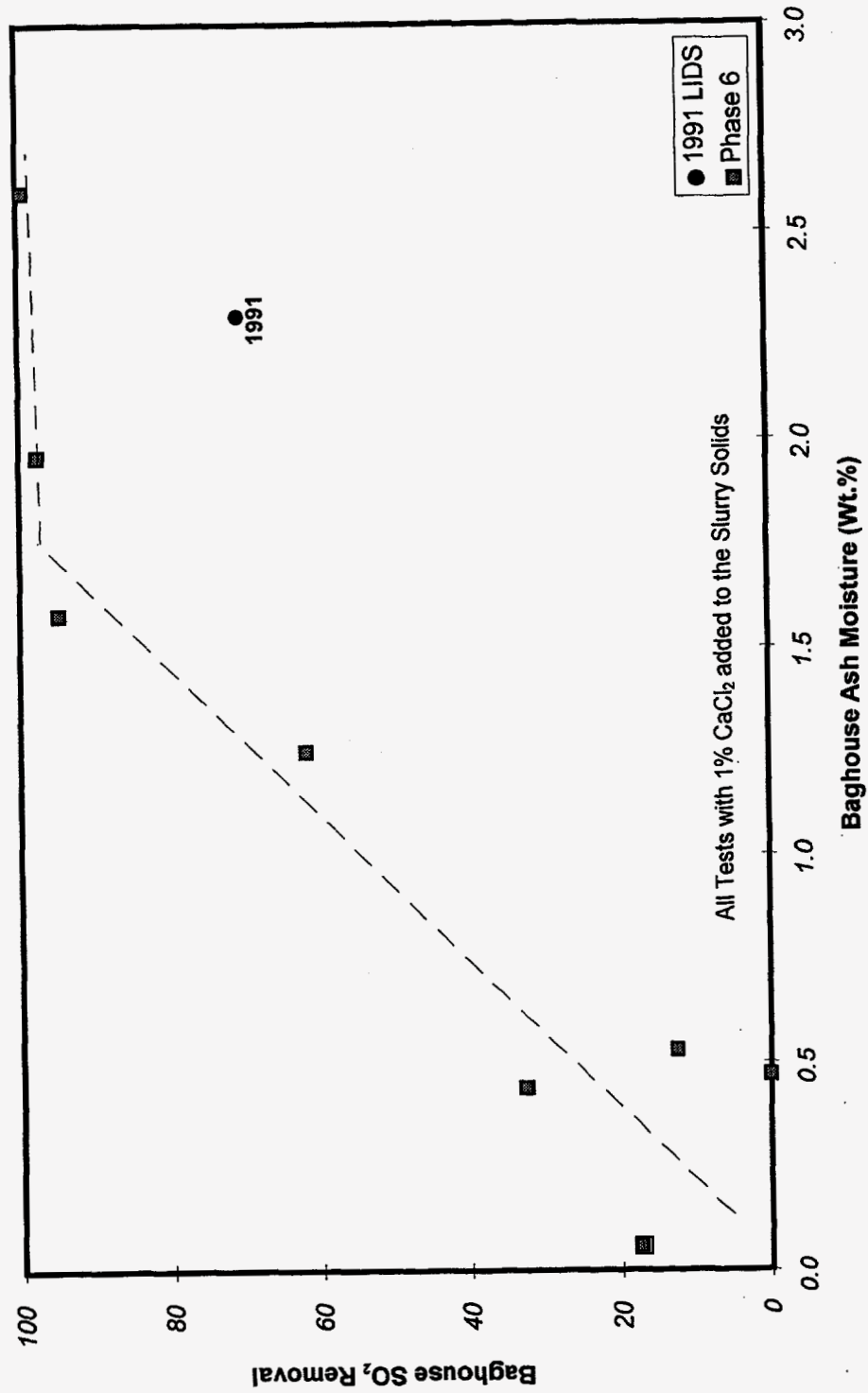
Total A+LIDS SO₂ Removal

Figure 18 shows total SO_2 removal as a function of T_{as}. Data are grouped by nominal furnace Ca/S ratios of 1.00, 1.22 and 1.44, and whether or not ammonia addition was employed. The figure shows that ammonia addition increases SO_2 removal in all cases. The greatest increase occurred at a furnace Ca/S ratio of 1.0 and a scrubber T_{as} of 10°F, when SO_2 removal increased from 87% to 99% with the use of ammonia. This exceeds the SO_2 performance goals for this phase and demonstrates the potential of A+LIDS for SO_2 control.

A+LIDS NO$_x$ Removal

Figure 19 shows furnace NO_x removal as a function of furnace NH_3/NO_x ratio. The maximum NO_x reduction was 56% at a NH_3/NO_x ratio of 3.2. Ammonia slip was maintained below 10 ppm for this test and below 15 ppm for all tests by limiting the amount of excess ammonia fed to the furnace. The performance goal of 60% NO_x removal was nearly met with ammonia injection alone. Combined with Low NO_x burners that can achieve 50% NO_x reduction, these technologies have the potential for high NO_x reduction.

Ammonia Recovery

A bench-scale test was performed on waste samples to characterize the ammonia recovery process. The samples were mixed with heated water and the ensuing reaction was studied at various temperatures and reaction times. The composition of the final product was compared to the original sample to calculate ammonia recovery efficiency. Ammonia recovery was found to be more efficient at high temperature. Figure 20 shows ammonia recovery from a sample reacted at various temperatures for five minutes. At 80°F, about 14% of the ammonia was recovered. This increased to 35% at 160°F. The reaction required about 60 minutes to reach completion. Figure 21 shows ammonia recovery at various intervals for three samples tested at 160°F for 60 minutes. Ammonia recovery averaged about 80% with a maximum of 94%. This shows that the goal of 90% ammonia recovery is feasible.

Post-test inspections revealed no significant deposition on the atomizer, scrubber walls, DID, flues, or in the baghouse. The baghouse ash remained dry and free-flowing for all tests. This demonstrates the potential of B&W's dry scrubber advancements in atomizer design and deposition control to enable future dry scrubbers to operate cleanly at very low T_{as}.

2.2.7   Phase 7 - Regenerable MgO-Based Dry Scrubbing

In this phase, the magnesium oxide-based (MgO) dry scrubbing process was investigated. MgO has been used commercially in a wet scrubbing process but has never been used in a dry scrubber. The advantage of an MgO-based system is that the reaction product, magnesium sulfite ($MgSO_3$), decomposes at much lower temperatures than calcium sulfite ($CaSO_3$) and can, therefore, be economically regenerated. Wet systems produce a sludge that must be dewatered and dehydrated before regeneration. Dry systems eliminate the dewatering step at a considerable energy savings.

A schematic of the regenerable MgO dry process is shown in Figure 22. This process is similar to conventional lime dry scrubbing, except the waste product can be economically regenerated so that only small quantities of makeup MgO are required. Since the MgO process is essentially a closed loop, it is necessary to remove fly ash from the flue gas prior to the dry scrubber so that it does not amass in the system. The process begins as dust-free flue gas enters the dry scrubber where it is

FIGURE 18   Phase 6 - Total SO₂ Removal
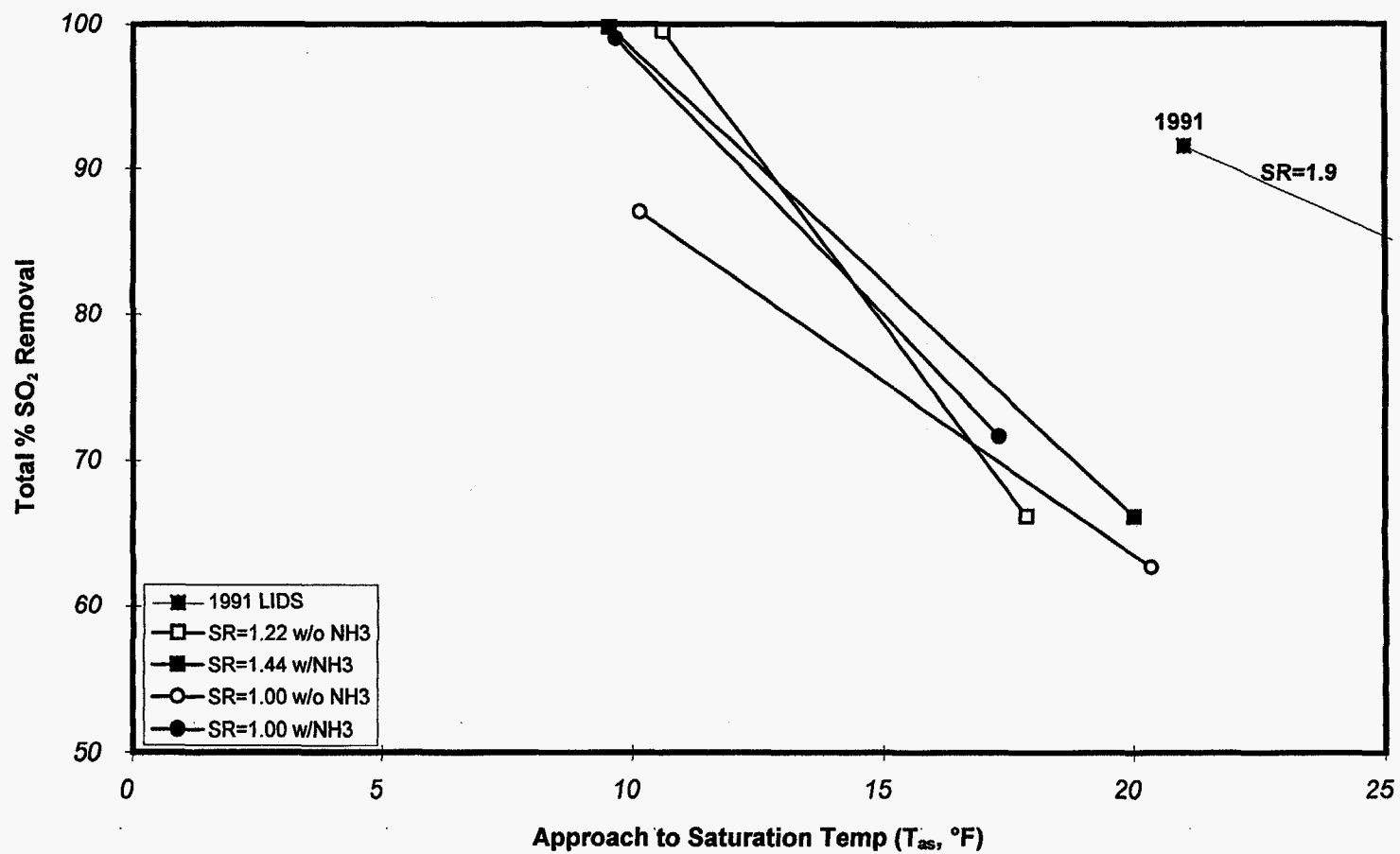
FIGURE 19 Phase 6 - Furnace NOₓ Removal
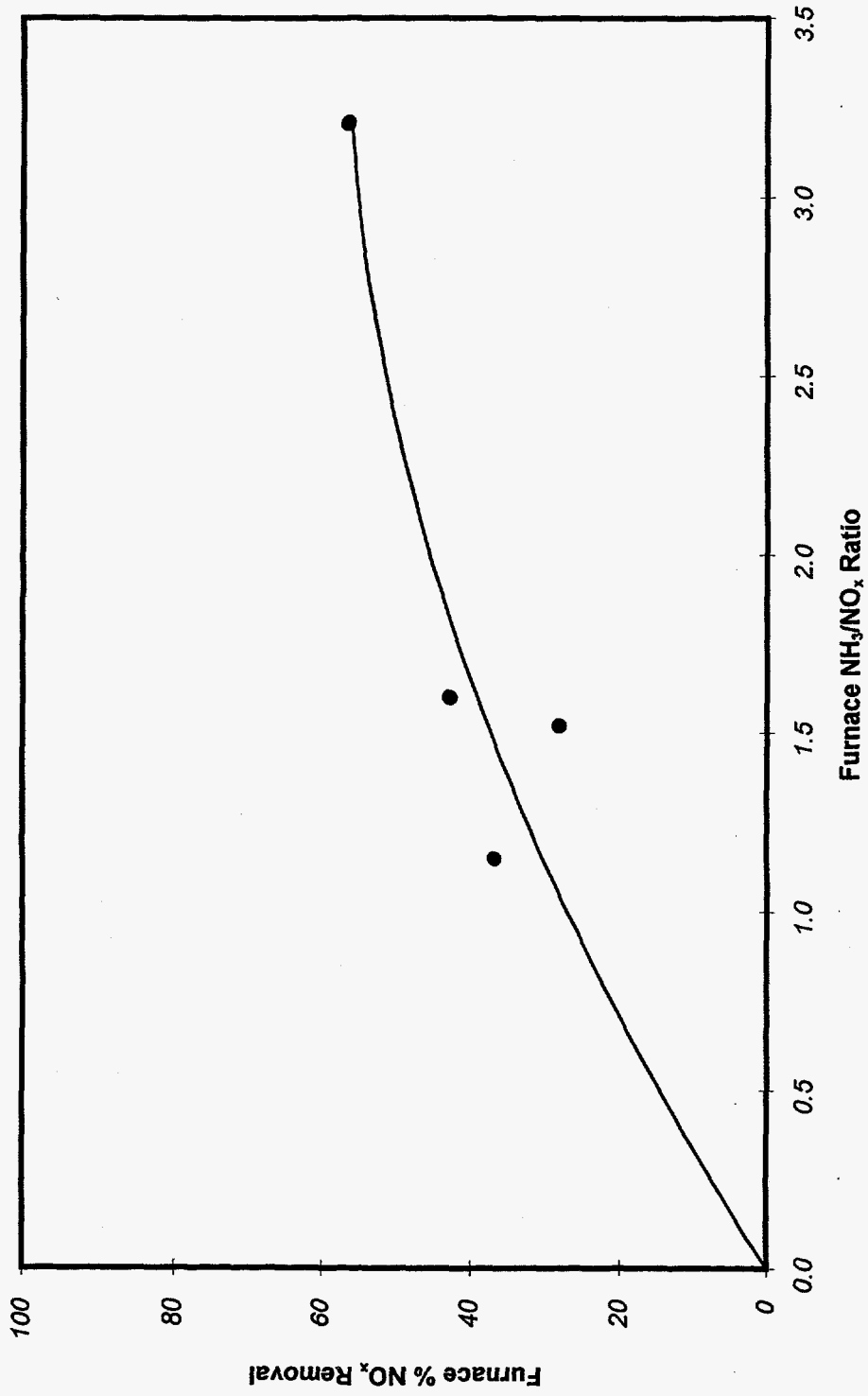
FIGURE 20   Ammonia Recovery - Effect of Temperature

Chart showing % NH₃ Recovery after 5 min. vs Reaction Temperature. Y-axis (Reaction Temperature): 160°F, 120°F, 80°F. X-axis (% NH₃ Recovery after 5 min.): 0 to 50. Bars: 160°F ≈ 35, 120°F ≈ 22, 80°F ≈ 14.
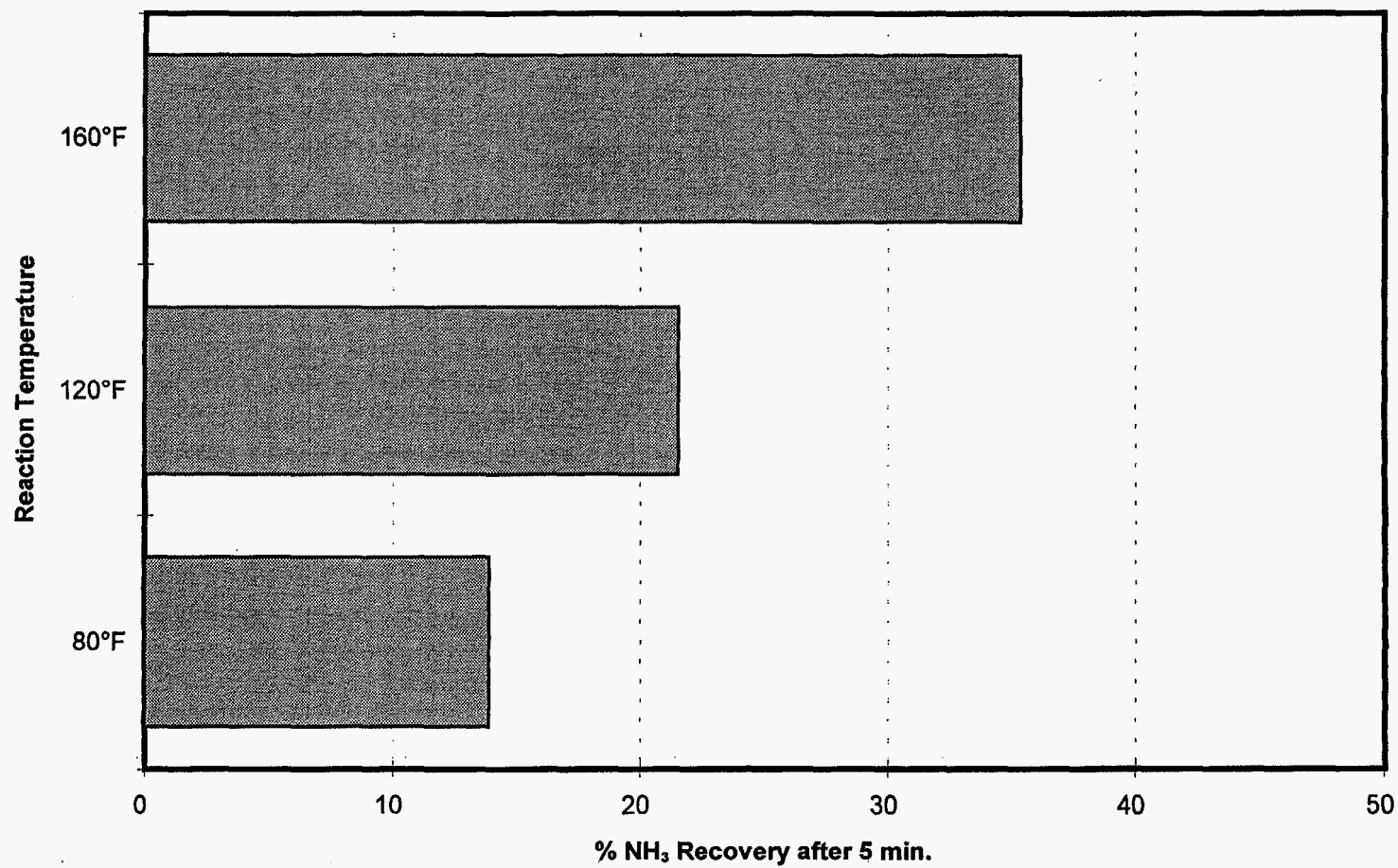
FIGURE 21 Ammonia Recovery - Yield at 160°F and 60 min.

Reaction Time @ 160°F (min.)

% NH₃ Recovery

Test 9
Test 10
Test 13

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FIGURE 22

Regenerable MgO Dry Scrubbing Process Schematic

Regenerable MgO Dry Scrubber

contacted with atomized droplets of magnesium sulfite slurry. The droplets absorb SO_2 from the flue gas and dry before exiting the scrubber. Dried particulate is collected in a second particulate collection system, transported to a regeneration system, and heated to drive off SO_2 and produce new MgO for the scrubber. The gas from the regenerator contains about 15% SO_2 and can be used to produce sulfuric acid or other sulfur by-products. An undesirable reaction that can occur in the scrubber or regenerator is the oxidation of the magnesium sulfite to sulfate. Magnesium sulfate has a higher decomposition temperature and is, therefore, more costly to regenerate. The same antioxidants used successfully in wet scrubbers were tested during this phase.

The important parameters in this process are the amount of reagent in the slurry, the amount of slurry injected into the scrubber (used to calculate the scrubber Mg/S ratio), the scrubber and baghouse T_{as}, the concentration and type of antioxidants, and the temperature and residence time in the sorbent regenerator.

The major chemical reactions that occur in the slaker, dry scrubber, baghouse, and regenerator are shown below. For greatest efficiency, the scrubber and baghouse SO_2 reactions must take place in the liquid phase.

Slaking Reaction:
$$MgO + H_2O \rightarrow Mg(OH)_2 + Heat$$

Scrubber and Baghouse Reactions:
$$SO_2 + MgSO_3 + H_2O \rightarrow Mg(HSO_3)_2$$
$$Mg(HSO_3)_2 + Mg(OH)_2 \rightarrow 2MgSO_3 + 2H_2O$$
$$Mg(HSO_3)_2 + MgO \rightarrow 2MgSO_3 + H_2O$$
$$MgSO_3 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2O$$
$$MgSO_3 + \frac{1}{2}O_2 \rightarrow MgSO_4 \quad \text{(undesirable oxidation reaction)}$$

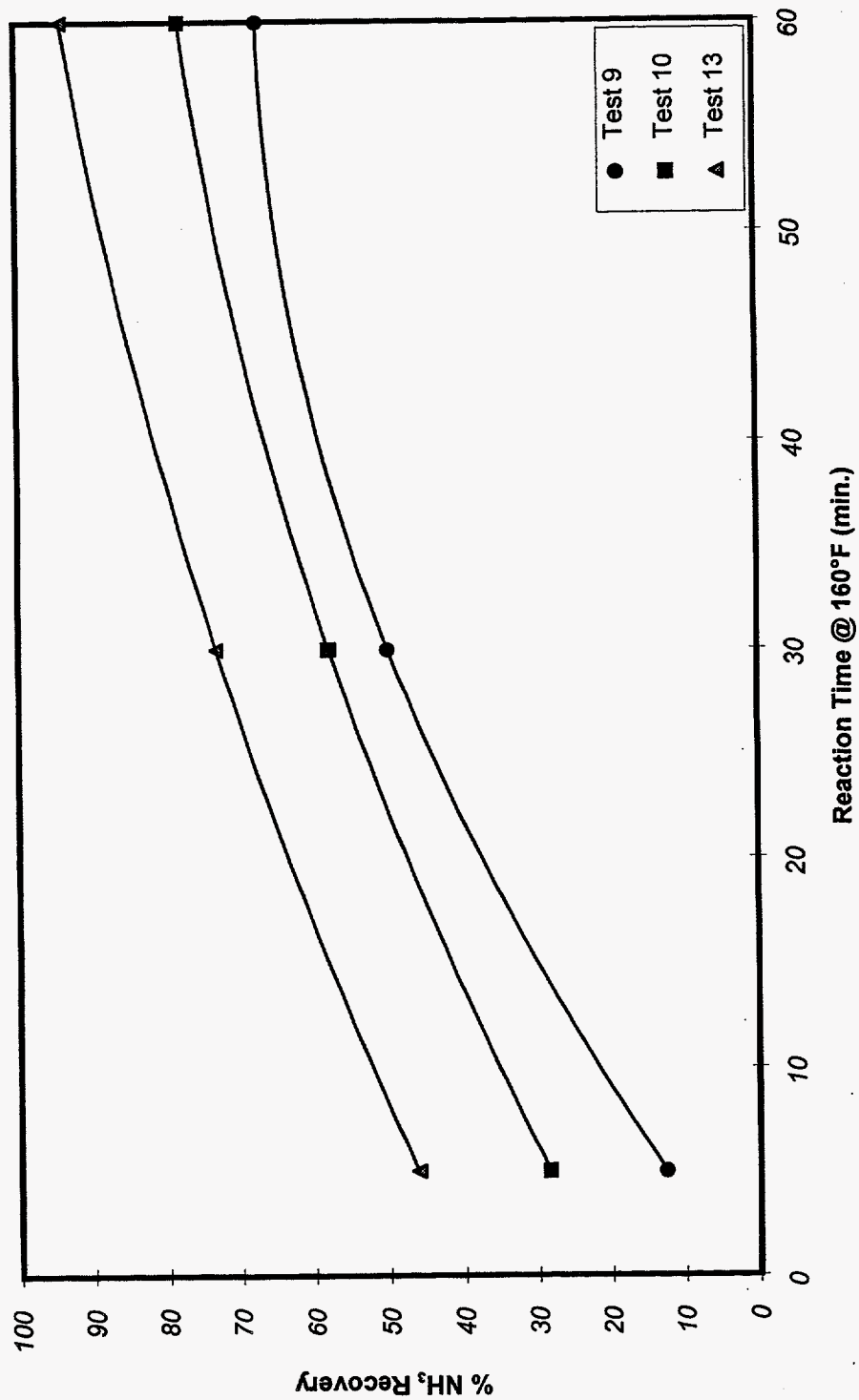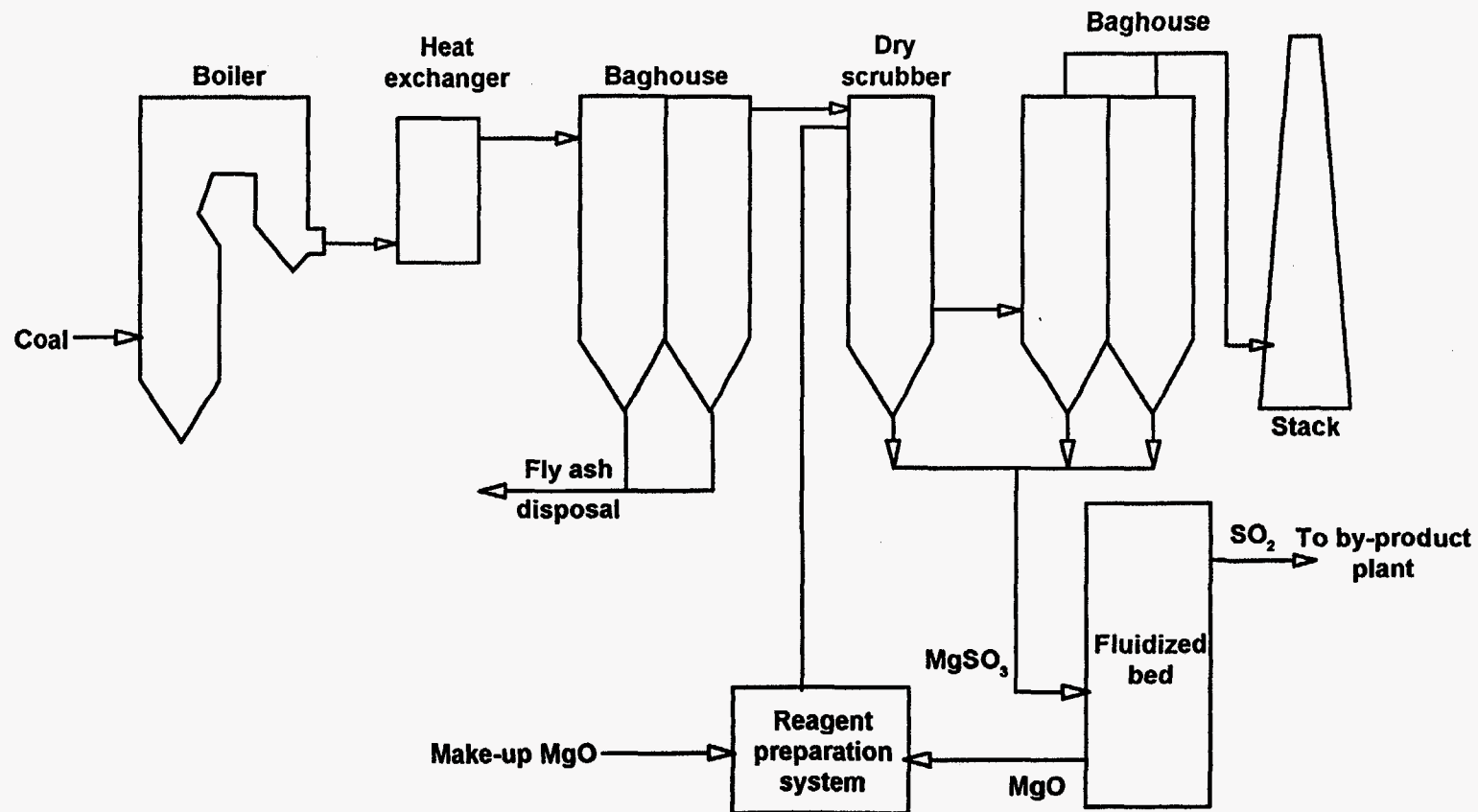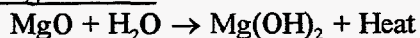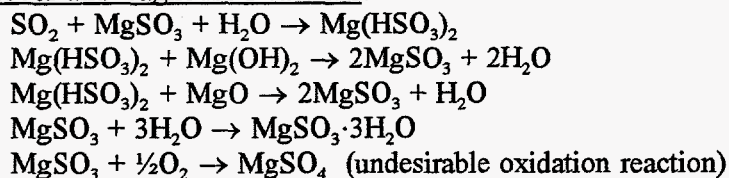Fluid Bed Regeneration (900°F):
$$MgSO_3 \cdot 3H_2O + Heat \rightarrow MgSO_3 + 3H_2O$$
$$MgSO_3 + Heat \rightarrow MgO + SO_2$$
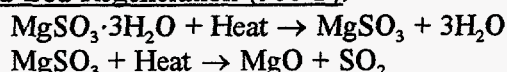
Although regenerable magnesia-based wet scrubbing has been commercially demonstrated, regenerable magnesia-based dry scrubbing has not. In a magnesia-based wet scrubber the primary reactions are:

$$MgSO_3 \cdot 6H_2O \rightarrow Mg^{2+} + SO_3^{2-} + 6H_2O$$
$$SO_3^{2-} + SO_2 + H_2O \rightarrow 2HSO_3^-$$
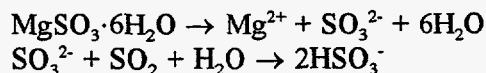
The role of $Mg(OH)_2$ is to convert bisulfite back to sulfite via:

$$Mg(OH)_2 \rightarrow Mg^{2+} + 2OH^-$$
$$OH^- + HSO_3^- \rightarrow SO_3^{2-} + H_2O$$
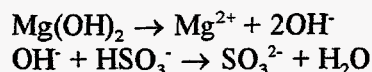
The important point is that in a magnesia system, $Mg(OH)_2$ does not react directly with SO_2. By contrast, SO_2 can react directly with $Ca(OH)_2$ in a lime-based scrubber. The difference between $Ca(OH)_2$ and $Mg(OH)_2$ is that the former is 20 to 100 times more soluble than $Mg(OH)_2$, depending on the temperature of the water. On the other hand, $MgSO_3 \cdot 6H_2O$ is 300 to 800 times more soluble than $CaSO_3 \cdot \frac{1}{2}H_2O$. Therefore, ash recycle is imperative in magnesia-based dry scrubbing to ensure that magnesium sulfite is present to react with SO_2.

2.2.8 Phase 7 - Results

The objective of this phase was to determine the feasibility of MgO dry scrubbing and fluidized-bed sorbent regeneration. The performance goal was 90-95% SO_2 removal and 90% MgO recovery. The furnace was modified to burn natural gas to simulate dust-free flue gas. An SO_2 injection system was used to spike the flue gas to the desired SO_2 concentration. Four tests were conducted at fresh Mg/S ratios ranging from 0.47 to 1.62 and T_{as} from 13 to 23°F. To simulate steady-state operation with recycle, $MgSO_3$ was added to the fresh slurry in the storage tank.

Table 8 is a summary of important operating parameters and performance results. The upper portion contains flue gas temperature and pressure drop for the major components, and atomizer slurry and air flows. The lower portion contains the fresh scrubber stoichiometry, scrubber T_{as}, flue gas flow rate at the scrubber inlet, SO_2 concentrations adjusted to 3% oxygen, SO_2 removal data, and sorbent utilization. A list of nomenclature is also provided.

Dry Scrubber SO_2 Removal

In Figure 23, SO_2 removal across only the dry scrubber is shown as a function of fresh Mg/S. Lime-based dry scrubbing data from Phase 4 is also included as a comparison. Removal is based on the scrubber inlet and outlet SO_2 concentration corrected to 3% oxygen. The average T_{as} for this phase was 17°F. Dry scrubber SO_2 removal ranged from 14 to 64% at Mg/S ratios of 0.47 to 1.63. The figure shows that this is about 15% lower than calcium-based systems. At a Mg/S ratio of 1.6 and a T_{as} of 15°F, the scrubber removed about 64% of the inlet SO_2 compared to 80% for a lime scrubber. However, no removal was observed across the baghouse and, therefore, the overall performance of this technology was only marginal. The cause of the poor performance in the baghouse is not completely understood and requires further investigation before MgO dry scrubbing becomes a viable technology. The bench-scale sorbent regeneration tests were postponed because of the poor overall performance.

There were two differences in the operation and performance of the dry scrubber during this phase compared to tests with lime slurry. To simulate dust-free flue gas, the pilot combustor was fired with natural gas, which resulted in a flue gas saturation temperature of 140°F compared to 120°F for coal. To compensate, the dry scrubber inlet temperature was increased by 20°F to 320°F. Also, periodic inspections of the DID showed significant amounts of wet deposits at a T_{as} of 10°F. This dictated that the scrubber T_{as} be limited to 15°F. Tests with lime slurry of similar solids concentration could be conducted at T_{as} less than 10°F with no DID deposits.

The most significant difference from a lime-based system was that no SO_2 removal occurred across baghouse. For lime systems, over 95% SO_2 removal can occur in the baghouse. Dry scrubber results from Phase 4 showed a strong relationship between SO_2 removal and T_{as}. Ash moisture and SO_2 removal increased as T_{as} decreased. For this phase, moisture in the baghouse solids ranged from 0.6% to 2.0%. Although this is similar to moisture levels from Phase 4, the difference in reaction chemistry may necessitate higher free moisture in the filter cake for MgO scrubbing. As stated previously, SO_2 does not react directly with the fresh sorbent, it reacts with $MgSO_3$ to form bisulfite, which then reacts in the aqueous phase with the fresh sorbent to form additional $MgSO_3$.

TABLE 8 Summary of Phase 7 - Regenerable MgO-Based Dry Scrubbing

Test	Test Description	Temperature (°F)			Differential Pressure (in.)			Atomizer		Atom Air	Vent Air	A:W	VA:AA
		DSin	DSout	Twb	HX	DS	BH	Slurry (lb/hr)	% Solids	Air (lb/hr)	Air (lb/hr)	(lb/lb)	(lb/lb)
1	SR=1.6, T_{as}=15F	321	161	144	9.1	2.3	3.0	204	26.6	127	154	0.62	1.21
2	SR=0.5, T_{as}=15F	320	157	142	9.2	2.5	2.6	219	14.2	97	114	0.44	1.18
3	SR=1.0, T_{as}=15F	323	157	143	9.6	2.5	2.2	213	20.0	93	114	0.44	1.22
4	SR=1.2, T_{as}=25F	325	169	145	9.7	2.5	2.4	170	23.2	92	114	0.54	1.24

Test	Test Description	SR Fresh Mg/S	T_{as} (°F)	DSin Gas (lb/hr)	SO$_2$ Data (ppm @ 3% O$_2$, dry)			SO$_2$ Removal (%)			Ca Utilization
					DSin	DSout	BHout	DS	BH	Total	
1	SR=1.6, T_{as}=15F	1.63	17	4111	2598	925	#N/A	64.4	#N/A	#N/A	#N/A
2	SR=0.5, T_{as}=15F	0.47	15	4172	2959	2529	#N/A	14.5	#N/A	#N/A	#N/A
3	SR=1.0, T_{as}=15F	1.03	14	4397	2989	1724	1836	42.3	-6.5	38.6	37.5
4	SR=1.2, T_{as}=25F	1.16	23	4311	2437	1359	1411	44.2	-3.8	42.1	36.2

Nomenclature

Twb = Wet Bulb Temperature

HX = Heat Exchanger

BH = Baghouse

DS = Dry Scrubber

T_{as} = Approach to Saturation Temperature
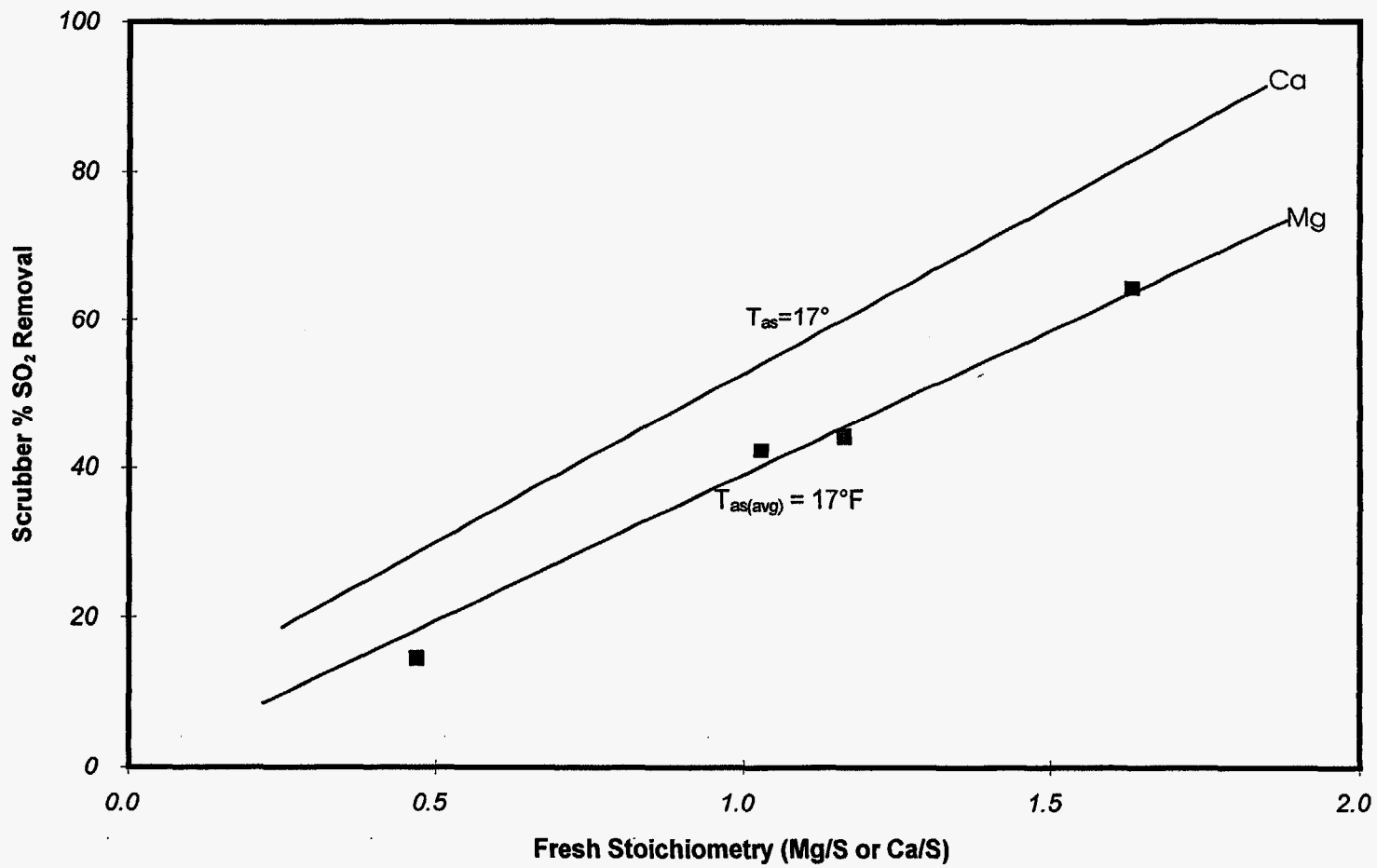
A:W = Air to Liquid Ratio of Atomizer

VA:AA = Ratio of Vent Air to Atomizing Air

SR = Magnesium to Sulfur Molar Ratio

GTA 1/20/95 43117 RED7.XLS

FIGURE 23   CaO- and MgO-Based Dry Scrubber SO₂ Removal

In the MgO process, it is also important to minimize the oxidation of magnesium sulfite because magnesium sulfate requires a higher regeneration temperature. During this phase, EDTA (160 ppm) and p-aminophenol (10 ppm) were used as antioxidants in the feed slurry. Ash analyses of two samples taken from the dry scrubber hopper showed that most of the sulfur occurred in the sulfite form. However, 12% oxidized to the sulfate indicating that the antioxidants were not completely effective.

System Operability

All tests were conducted with the DID installed at the scrubber outlet. The DID performed well throughout this phase. In previous phases using lime slurry above 10% solids, little material collected on the DID. During this phase, however, it was noticed that deposition on the DID was greater, even at approach temperatures of 25°F and solids concentrations of 20%. In addition, the deposits had a greater tendency to remain moist at low T_{as}. Consequently, T_{as} was limited to 15°F.

The texture of the DID deposits also differed from lime-based deposits in that they tended to be more granular. Ash from the baghouse hopper was also more granular, but still flowed freely and caused no handling problems. The granular nature of the ash raises concerns over their suitability as a recycle material. In a commercial application, the large granules would have to be broken up or strained to prevent plugging the atomizer and interfering with atomization quality.

Post-test inspections revealed no significant deposition on the atomizer, scrubber walls, flues, or in the baghouse. At a T_{as} of 15°F, baghouse ash appeared dry and flowed freely. As mentioned above, deposition on the DID increased for this phase, but was compensated for by operating at higher T_{as}.

2.2.9   Project Conclusions and Accomplishments

SO_2/NO_x Removal

The most significant highlight of this project is that the SO_2 removal goals were surpassed for all of the lime-based technologies. This demonstrates ability of these technologies to minimize pollution when burning high-sulfur, Eastern coal.

Results of Ca-Based Advanced Dry Scrubbing

Technology	Ca/S Ratio	T_{as}, (°F)	% SO_2 Removal
Dry Scrubbing	1.15	10	98
LIDS	1.44	11	99
A+LIDS - SO_2	1.00	10	99
- NO_x	3.2 NH_3/NO_x		56% NO_x

Low T_{as} Operation

The success of the new atomizer and DID allowed trouble-free operation at scrubber and baghouse approach temperature of 10°F and less. The nature of the deposits that formed on the DID was found to a function of solids concentration as well as T_{as}. Tests with lime-slurry produced no significant deposition at a T_{as} of 10°F as long as the solids concentration was greater than about 10%. The success of this project in attaining very high SO_2 removal is directly attributable to the ability to operate the system at low T_{as}.

Droplet Impingement Device (DID)

Most tests were conducted with the B&W patented Droplet Impingement Device (DID) installed at the scrubber outlet. Its purpose was to prevent deposition downstream of the scrubber by selectively collecting large, heavy, unevaporated slurry droplets, that may be present at low T_{as}. Smaller, lighter and dry particles have much less momentum and follow the gas stream around the DID without being collected. Provisions were made to clean the DID by rapping, rotating, or both. The DID performed well during all test phases. It was found to be needed only when the scrubber was operated at very low T_{as} with slurries containing less then 10% solids. Under these conditions, incomplete evaporation and low initial solids concentrations resulted in large, wet, slurry droplets at the scrubber exit. These droplets impinged the DID and gradually built deposits that had to be removed by rotating and rapping the DID. In a few cases, the deposits were moist and difficult to remove. However, even for these cases, the downstream flues and equipment remained deposit-free, demonstrating that the DID concept is viable during extreme or upset conditions. In tests using slurries above 10% solids, little material collected on the DID. The success of the DID throughout this test has lead to a study to develop a commercial system.

Atomizer Deposition

A proprietary atomizer was used throughout the project. Its design minimized deposition on the atomizer tip, which is a serious drawback to commercial dry scrubber systems. Deposition eventually affects atomizer performance which leads to reduced scrubbing performance, catastrophic deposition on scrubber walls, flues, and in the baghouse, and downtime for maintenance. The new B&W Durajet^{tm} prevented all but very slight deposition that was easily removed by light rapping every eight hours. It is hoped to incorporate this design on the next commercial dry scrubber or to retrofit existing scrubbers.

Effect of Moisture

Initial LIDS testing produced SO_2 removals of only 54%, compared to 92% in 1991. Sample analyses showed that the poor performance was likely caused by exceptionally fast evaporation in the scrubber as evidenced by the low moisture content of the baghouse ash and poor scrubber performance. At similar conditions, baghouse ash moisture was only 0.1% compared to 2.3% in 1991. To compensate, small amounts of a deliquescent salt ($CaCl_2$) were added to the slurry to increase the residual ash moisture in the baghouse. Tests without salt addition produced moisture levels below 0.5% and baghouse SO_2 removal below 40%. With the addition of $CaCl_2$, the residual moisture increased to 1.6% and SO_2 removal increased to above 80%. This effect is usually not seen, or is not as great, with conventional dry scrubbing systems, because it is believed that the chemical makeup of the solids (i.e. higher $Ca(OH)_2$ content) maintains residual ash moisture above 1.0%. Residual

baghouse ash moisture may be a new and simple method of diagnosing poor baghouse and dry scrubber performance. It's effect on conventional lime dry scrubbing warrants further research.

Baghouse Bag Material

Two new types of Teflon-treated baghouse bag materials were tested, Tuflex and Microtex. Both materials performed well throughout testing. No problems were encountered with particulate handling even at T_{as} as low as 8°F. The bags remained dry and deposit-free, and the ash flowed freely from the hoppers. Both materials perform equally well and much better than the felted Nomex used previously. This type of material will be recommended for use on any future B&W dry scrubber contracts.

2.2.10 Waste Characterization, Disposal, and Utilization

B&W recently completed an FGD waste characterization study[9] as part of the DOE Combustion 2000 Program. Sections of the study and other reports used in the following discussion were conducted by EERC, EPRI[10,11], The Ohio State University[12], and B&W[13]. Much of the study is reproduced below along with waste analyses from this project.

Characterization

Coal-fired utilities generate large amounts of solid by-products FGD technologies used to control the emission of sulfur dioxide. Due to future demand, and as more plants comply with Titles III and IV of the 1990 Clean Air Act Amendments, the amount of by-products produced are expected to increase. In the past, solid by-product generation has been a low priority issue. However, due to ever increasing public and governmental pressure, it has become important to assess the characteristics of these by-products in terms of their disposal and utilization characteristics.

As with many FGD processes, LIDS, A+LIDS, and dry scrubbing produce a dry by-product that contains mostly calcium oxide, calcium hydroxide, calcium sulfite, calcium sulfate, and coal flyash. These constituents have the potential to cause disposal problems, such as fugitive dust, and leachate with a high pH. Under RCRA, solid wastes are classified as hazardous and unsuitable for conventional disposal if they exhibit characteristics of ignitability, corrosivity, reactivity, and/or extraction procedure toxicity as defined by RCRA The main environmental concern associated with coal combustion by-products is the potential leaching of harmful constituents into groundwater.

In 1988, EPA submitted a report to Congress summarizing its analysis of the potential effects of the coal-use residues. The report indicated that the wastes were generally non-hazardous and concluded that existing waste management practices and state regulatory requirements were adequate for protecting human health and the environment. However, the EPA did not make a final regulatory determination for fly ash, bottom ash, boiler slag, and flue gas emission control waste until August 9, 1993. At that time the EPA concluded that the hazardous waste exemption for these wastes should continue. Currently, coal-combustion wastes are regulated under RCRA Subtitle D non-hazardous waste regulations which grant regulatory authority to the States.

EPA Hazardous Waste Characterization[13]: LIDS by-product was analyzed for ignitability, corrosivity by pH, reactive sulfide, and reactive cyanide using U.S. EPA Methods. Method 1311 (TCLP) was used to characterize leachates for organics, volatile organics, and heavy metals. Table 9 shows that all levels were below hazardous waste guidelines, and that all organic, volatile organic, and metal analyses were below detection limits and TCLP limits.

Ohio Water Quality Standards[13]: The regulatory status of clean coal by-products in the State of Ohio in terms of water quality standards are as follows:

> Clean coal combustion by-products are considered solid waste under the jurisdiction of the Ohio EPA, Division of Solids and Hazardous Waste, with exemptions awarded on a case-by-case basis...

> Ohio has a general exemption for nontoxic fly ash, which is not regulated as a solid waste; however, disposal of this by-product is subject to Ohio water pollution and air pollution regulations under the jurisdiction of the Ohio EPA, Division of Water Pollution Control. Utility fly ash is considered to be nontoxic and, therefore, is exempt from regulation as a hazardous or solid waste if its leachate does not exceed 30 times the levels specified in the Ohio Drinking Water Standards (Ohio Revised Code 3745-81-11(B)).

Ohio water quality standards for metals were compared to the leachate from LIDS samples. Table 10 shows that metal concentrations in 20:1 water leachate samples were well under Ohio and Federal guidelines.

Ohio State University (OSU) Study[12]: OSU recently completed Phase I of a study to determine if FGD by-product can be used as a substitute for materials now being used for land reclamation. The work was performed with cooperation from the United States Geological Survey and Dravo Lime Company.

A total of 58 FGD samples from LIDS, Spray Drying, Duct Injection, LIMB, Coolside, fluidized bed combustion, and pressurized fluidized bed combustion were analyzed. Wide variations in elemental composition appeared to be more a function of variations in the coal composition than in the type of clean coal technology. Results of the by-product characterization are as follows:

> Many of the elements regulated by the U.S. EPA reside primarily in the coal fly ash. The presence of excess sorbent and reaction products was found to cause a dilution of the concentration of these elements in the by-product materials as compared to fly ash alone.

> The leachate pH for some samples investigated were near the RCRA limit of 12.5 for toxic waste, however none surpassed the limit. "The leachates were obtained on fresh samples, however, upon aging and exposure to moisture and atmospheric or soil CO_2, the CaO and $Ca(OH)_2$ will carbonate to form calcite ($CaCO_3$) and the pH will decrease to around 8.3."

> Concentrations of the eight RCRA metals (Ag, As, Ba, Cd, Cr, Hg, Pb, and Se) were below drinking water standards for both ASTM and TCLP leachates.

TABLE 9   Hazardous Waste Characterization for LIDS

Test	EPA Method No.	LIDS By-Product	EPA Hazardous Waste Limit
Ignitability	1010	>150	
Corrosivity (pH)	1110	12.34	12.5
Reactive Sulfide	9040	<48	500
Reactive Cyanide	9040	<35	250
TCLP Metals	1311	All Below Regulatory Limits	
TCLP Semivolatile Organics	1311	All Below Regulatory Limits	
TCLP Volatile Organics	1311	All Below Regulatory Limits	
Source: "Management of Solid Wastes from the LIDS Clean Coal Technology"3, Czuzwa, Musiol			

TABLE 10   Ohio Water Quality Standards for Metals

Parameter	OH Drinking Water Standards mg/l	Federal/State 30 x Standard mg/l	RCRA Criteria mg/l	LIDS Water Leachate mg/l
Arsenic	0.05	1.5	5.0	<1*
Barium	1.0	30	100.0	<0.5
Cadmium	0.01	0.3	1.0	0.004
Chromium	0.05	1.5	5.0	0.083
Lead	0.05	1.5	5.0	0.012
Mercury	0.002	0.06	0.2	<0.04*
Selenium	0.01	0.3	1.0	<0.1*
*not measured in the ASTM water leachate; value given is for the TCLP acetic acid leachate.				
Source: "Management of Solid Wastes from the LIDS Clean Coal Technoloty"3, Czuzwa, Musiol				

Particle size analysis indicated that all fly ash materials were very fine with those from the duct injection, spray dryer, and lime injection process having in excess of 80% by weight finer than 0.025 mm effective diameter. The particles from all of the processes, with the exception of the fluidized bed fly ash, were quite uniform. This is relevant because it defines the type and size of equipment used to handle a product and the behavior of the material during handling.

All of the by-products would be considered light weight materials -- compacted densities lower than those of a typical natural soil. The volume occupied by a given weight of end-product material will be a factor in sizing, product handling, storage and transportation equipment, and in the disposal volume requirements.

EPRI Advanced SO$_2$ Control Solid-Waste Management Planning Study[10]: This study summarized available information on clean coal and wet FGD waste properties that affect how these materials are disposed or reused. The following statement is from the report:

> "Solid waste from spray dryer systems can be disposed of in landfill operations similar to those used for wet-scrubber wastes which have been blended with fly ash and lime. Because the waste material from the spray dry process is fine and dry (1-5% moisture), water is normally added at the landfill for dust control. The addition of water to 10-30%, can cause pozzolanic reactions to take place between the mixture components resulting in a waste material which normally has an extremely low permeability when compacted and desirable landfill characteristics, such as increased compressive strength and reduced ion leaching rates."

It can be assumed that these desirable landfill characteristics will be present in the by-products produced during this project because of their similarity to spray dryer by-product,

Waste Characteristics of this Project: Table 11 summarizes the analyses performed on waste and slurry samples from this project. It also shows the amount and major constituents of typical wastes produced for Phases 4, 5 and 6. The main difference between the by-products described so far and those from this project, is that this project produced material with lower concentrations of unused sorbent (Ca(OH)$_2$ and CaO). This is because the technology advancements employed throughout this project permitted scrubber operation at very low T_{as}. This greatly increased scrubber and baghouse efficiency and sorbent utilization, and decreased sorbent usage and waste production. Several tests achieved sorbent utilization above 90%, and the amount of reactive sorbent remaining in several samples was so low, that the corresponding leachate had pH values below 10.

Waste Utilization and Disposal

Utilization is an attractive economic and environmental alternative for managing the high volume of waste generated by FGD processes. Clean coal technology by-products exhibit significantly different characteristics from conventional coal by-products such as fly ash, bottom ash, and boiler slag, because they contain high concentrations of sorbent-derived components. Conventional boiler wastes have been utilized as engineering and construction materials in numerous applications, and alternate uses continue to be researched and developed. High-calcium and high-sulfur wastes exhibit different utilization characteristics, and so their potential applications and markets are also different. Utilization potential for FGD wastes is also affected by coal composition, boiler type and operating conditions, fly ash and sorbent composition, and the relative amounts of fly ash and unreacted and spent sorbent.

TABLE 11   Waste Characterization for Phases 4, 5, and 6

Chemical Analyses

	Phase 4B		Phase 5			Phase 7			
Recycle Slurry	7	8	D7	D8	D10	7	9	10	13
% Solids @ 45 C	38.6	39.5	37.9	38.5	33.2	50.2	46.9	45.9	44.2
AEHL, % $Ca(OH)_2$	3.1	2.6	27.4	28.3	22.4	5.7	3.5	0.2	0.1
% CO_2			6.7	5.5	5.9	4.5	5.9	3.5	3.0
Major Constituents by ICP									
Silicon, % SiO_2	15.3	14.7	8.5	8.5	9.1	11.5	9.1	10.2	10.9
Aluminum, %Al_2O_3	9.2	9.2	5.5	5.4	5.6	6.5	5.1	5.6	5.6
Iron, % Fe_2O_3	7.8	7.6	6.8	6.3	6.8	6.5	5.3	5.8	5.9
Titanium, % TiO_2	0.4	0.3	0.2	0.2	0.2	0.3	0.2	0.2	0.3
Calcium, % CaO	29.0	29.4	44.1	43.6	36.2	30.2	28.5	30.6	30.7
Magnesium, % MgO	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.6	0.6
Total Sulfur, % SO_3	34.4	35.4	26.0	28.4	34.4	26.2	28.5	39.0	37.7
Sulfate Sulfur, %SO_3	4.8	3.2	12.5	11.9	11.9	10.8	9.2	10.8	11.5
Sulfite Sulfur, % SO_3	29.6	32.2	13.5	23.6	22.5	15.4	19.3	28.2	26.2
Phosphorus, % P_2O_5	<0.50	<0.50	<0.5	<0.5	<0.5	<1.0	<1.0	<1.0	<1.0
Sodium, Na_2O	0.1	0.1	0.2	0.2	0.2				
Potassium, K_2O	0.6	0.5	0.4	0.4	0.4				
Baghouse Ash									
% H_2O @ 45 C	1.2	1.1	1.7	1.7	1.6	0.5	2.6	1.6	0.1
% Cl in Ash		0.9	1.4	1.4	1.4	1.8	1.7	1.4	1.2
Lime Slurry									
% Solids @ 105 C	17.8	18.1							
ALI, Wt.% $Ca(OH)_2$	93.3	92.5							

Waste Composition

Test Number	7	8	D7	D8	D10	7	9	10	13
Production, lb/hr	132				120			106	
Composition (wt.%):									
$CaSO_3$*1/2H_2O	47.7	51.9	21.7	26.6	36.3	24.8	31.1	45.5	42.3
$CaSO_4$*2H_2O	10.3	6.8	27.0	25.5	25.6	18.4	15.7	18.4	19.5
$Ca(OH)_2$	3.1	2.6	22.9	22.1	6.1	8.1	1.3	0.0	0.7
$CaCO_3$	0.0	0.0	15.3	12.6	13.5	10.1	13.4	8.0	6.9
Fly Ash	47.5	46.4	28.1	28.0	34.2	33.3	33.0	43.1	42.0
Total	108.6	107.7	115.0	114.8	115.6	94.8	94.5	114.9	111.3

2-53

A by-product utilization study[9] was recently completed to identify ways to reduce waste through alternate methods of utilization. The study followed the developments of several independent solid waste management programs to take advantage of knowledge gained. The Energy and Environmental Research Center (University of North Dakota) acted as a consultant on the study and supplied the following summary:

LIDS waste was characterized using numerous standard test procedures to evaluate its composition and behavior[13]. The results of these tests may be used to make a preliminary evaluation of its utilization potential. The primary areas of interest; chemical composition, mineralogical composition, and the physical/engineering characteristics, are discussed below.

Chemical Composition - The bulk chemical composition of coal combustion by-products has traditionally focused on major components and has reported these components as common oxides. Table 12 shows the range of the major constituents for several clean coal wastes and for typical fly ash. The information clearly shows the difference between conventional fly ash and clean coal by-products. It also shows that LIDS waste is similar to other clean coal wastes, especially LIMB and spray drying. The pH of the LIDS waste is high, as is typical of many clean coal technology residues, which may be advantageous in several utilization applications.

Mineral Composition - The bulk chemical composition shows only what elements are present in a material. The mineralogical composition tells how these elements are combined. Mineralogy affects how a material can be utilized or disposed. Table 13 is a comparison of the mineral phases found FGD wastes and fly ash. Analyses have shown the presence of hannebachite ($CaSO_3 \cdot \frac{1}{2}H_2O$), gypsum ($CaSO_4 \cdot 2H_2O$), calcite ($CaCO_3$), portlandite ($Ca(OH)_2$), lime ($CaO$), and fly ash. Hannebachite may also oxidize when exposed to the atmosphere or when aged to form gypsum.

Quartz, mullite, spinel, hematite, periclase, melilite, and merwinite are generally nonreactive minerals and do not contribute significantly to the cementitious or pozzolanic reactions that occur when these materials are hydrated for utilization or disposal. The minerals that contain Ca (hannebachite, lime, gypsum, anhydrite, and portlandite) are reactive when exposed to water. LIDS analyses show the presence of hannebachite, gypsum, and portlandite to be similar with spray drier wastes.

Physical/Engineering Characterization - The physical characteristics of a waste depend upon its solid-phase characteristics, before and after environmental interactions mineralogically alter the original materials. Physical characteristics are significant because they relate to hardness, compressive strength, flexural strength, particle-size distribution, ductility, permeability, bulk density, material handling, transportation, disposal, and utilization. The results of several studies investigating the physical characterization of clean coal wastes and fly ash are summarized in Table 14. LIDS wastes were found to be comparable to other clean coal wastes, indicating similar utilization potential.

Waste Utilization Alternatives - LIDS wastes are similar to many other clean coal by-products such as limestone injection, spray drying, duct injection, and fluidized-bed combustion. The table below summarizes alternative uses for several clean coal technology by-products. These applications are prioritized as having a high, medium, or low potential. In the case of the fluidized-bed combustion, the values encompass a wide range of FBC wastes.

TABLE 12 Characterization of Clean Coal Technology Wastes

Element	LIDS		Duct Injection		LIMB		FBC		Spray Dryer		Fly Ash	
	Low	High	Low	High	Low	High	Low	High	Low	High	Low	High
SiO_2	4.71	10.3	13.6	38.9	14.1	41.3	4.28	7.27	4.7	29.3	19.8	66.5
Al_2O_3	1.89	5.48	3.2	19.3	6.6	20.6	1.13	5.29	1.89	16.3	3.06	33.6
Fe_2O_3	2.15	3.58	4.94	10.4	7.15	12.7	1	3.86	2.14	7.29	2	26.2
CaO	34.4	56.1	16.4	48.2	14.98	50.4	33.7	57.7	24.92	56.14	0.64	50.1
MgO	1.25	25.1	0.49	1.11	0.65	21	0.63	26.9	0.06	25.12	0.5	8.8
Na_2O	--	--	0.11	4.27	--	--						
K_2O	0.23	0.55	0.31	1.7	0.48	2.23	0.1	0.18	0.19	0.6	0.04	3.56
TiO_2	--	--	0.09	0.59	--							
P_2O_5	0.055	0.094	0.04	0.17	0.0153	0.136	0.0059	0.024	0.002	0.175	0.05	1.21
SO_3	19.3	25.5	6	24.5	6.75	15	21.3	35.5	14.5	42.5	0.04	11.6
S	7.7	10.2	2.4	9.8	2.7	6	8.5	14.2	5.8	17	0.016	4.64

TABLE 13   Mineralogy of Clean Coal Technology Wastes

LIDS	LIMB	Spray Dryer	Duct Injection	FBC	PC Fly Ash
Portlandite	Calcite	Hannebachite	Portlandite	Anhydrite	Quartz
Anhydrite	Anhydrite	Portlandite	Hannebachite	Lime	Mullite
Calcite	Lime	Quartz	Quartz	Portlandite	C3A
Lime	Quartz		Calcite	Calcite	Melilite
Gypsum	Mullite		Mullite	Quartz	Spinel
	Hematite		Hematite	Periclase	Hematite
				Hematite	Lime
					Anhydrite
					Merwinite
					Periclase

TABLE 14 Physical Characteristics of Clean Coal Technology Wastes

Property	LIDS	LIMB	PC Fly Ash	Spray Dryer	Duct Injection	FBC
Bulk Density Aerated, pcf		35-65	35-65	36-49	25-33	28-87
Bulk Density Settled, pcf		46-70	56-89	50-80	35-48	
Optimum Moisture, %	41	36-52	10-31	16-38	29-45	17-42
Maximum Dry Density, pcf	67.8	68-80	81-88	71-104	67-83	43-94
Mean Particle Size, mm	0.013-0.018	0.018-0.027	0.001-0.40	0.016-0.045		0.01-0.65
Specific Gravity	2.6-2.7	2.6-3.05	1.5-3.1	2.29-2.80	2.1-2.35	2-3
Permeability, cm/s	1.7×10^{-5}	$10^{-9} - 10^{-4}$	$10^{-7} - 10^{-4}$	$10^{-9} - 10^{-7}$	$10^{-6} - 10^{-5}$	$10^{-7} - 10^{-4}$
Specific Surface Area, m^2/s	3.8	2.8-3.93	0.2-3.1	1.6-2.5	8.7-13.0	2.8-16
Unconfined Compressive Strength, psi		50-500	50-300	100-900	30-50	300-1200

When coal-conversion residues are used as a borrow material in construction embankments and load-bearing structural fills, they have two major advantages when compared to most natural solids and rocks: their availability in urban areas and their light unit weight. Advanced clean coal residuals, such as LIDS, are even more lightweight than pulverized coal fly ashes currently being used in these applications. The low unit weight of coal combustion residues reduces the load on weak foundation soils. The low density allows the use of a smaller tonnage of material for a given volume of fill, thus reducing trucking costs.

Alternative Uses for Clean Coal Technology By-products[11]

Atmospheric Fluidized-Bed Combustion

High Potential	Moderate Potential	Low Potential
Soil Stabilization	Structural Fill	Cement Production
FGD Sludge Stabilization	Stabilized Road Base	Cement Replacement
Agricultural Use	Concrete Products	Concrete Blocks
	Sewage Sludge Stabilization	Gypsum/Wallboard
	Industrial Sludge Stabilization	Liner Material
	Synthetic Aggregate	Mineral Filler
	Lightweight Aggregate	
	Ceramic Products	
	Mineral Wool	
	Grout/Mine Backfill	
	Brick Production	

Spray Dryer

High Potential	Moderate Potential	Low Potential
Structural Fill	Cement Production	Gypsum/Wallboard
Grout/Mine Backfill	Cement Replacement	Metals Extraction
Stabilized Road Base	Soil Stabilization	
Synthetic Aggregate	Sludge Stabilization	
Lightweight Aggregate	Mineral Filler	
Mineral Wool	Agricultural Use	
Brick Production	Ceramic Products	
	Liner Material	

Limestone Furnace Injection

High Potential	Moderate Potential	Low Potential
Structural Fill	Cement Production	Gypsum/Wallboard
Stabilized Road Base	Concrete Block	Metals Extraction
Synthetic Aggregate	Soil Stabilization	
Mineral Wool	Sludge Stabilization	
Soil Stabilization	Mineral Filler	
Lightweight Aggregate	Agricultural Use	
	Lightweight Aggregate	
	Ceramic Products	
	Brick Production	
	Grout/Mine Backfill	
	Liner Material	

Conclusions

Based on the list above, and on pilot- and full-scale demonstration projects for similar clean coal technologies, the following alternative uses have the highest potential:

- Abatement of acid mine drainage and mine reclamation.
- Structural fill and other fills, including controlled low-strength material.
- Soil amendment (road base, subbase, and agriculture).
- Synthetic/lightweight aggregate
- Mineral Wool

2.3 Economic Analysis and Commercialization

Within the last few decades wet scrubbing with lime or limestone slurries has become the dominant commercial FGD technology. Worldwide, there are currently 581 FGD systems operating on a total capacity of about 150 GW_e[3]. Approximately 70% of the units, representing 124 GW_e of capacity, are based on lime- or limestone-wet scrubbing. About 20% of the units, or about 15 GW_e, utilize either sodium-based or lime-slurry dry scrubbing. The remaining 10% of the units use various regenerable processes of sorbent-injection technologies. Another 180 units, representing about 85 GW_e are planned or under construction, and will employ these technologies in approximately the same proportions. B&W will continue to develop and aggressively market two of the technologies studied during this project; Advanced Dry Scrubbing, and LIDS. Magnesium-based dry scrubbing will not be pursued because of the marginal performance demonstrated during the feasibility study. Funding may be sought to investigate, at bench-scale, the cause of the poor baghouse performance with MgO-based systems. A+LIDS will be marketed as a niche technology when additional NO_x removal is needed on facilities seeking SO_2 control technology. A+LIDS will also be offered as a means to polish SO_2 performance, while boosting NO_x removal, on future LIDS applications.

Advanced Dry Scrubbing

The technological advancements in atomizer design and deposition control demonstrated during this project will be incorporated into the next generation of B&W dry scrubbers. However, dry scrubbing is perceived by the market to be applicable only to low-sulfur coal applications, and a fear is prevalent in the industry that they cannot be operated below a T_{as} of 30°F without catastrophic deposition. This fear developed from early bad experiences during dry scrubber start-ups, and little has been done since to achieve lower T_{as} at commercial scale. However, research continued to improve our understanding of basic dry scrubbing concepts, and great strides have been made in flow-field modeling, atomizer design, and deposition control. This project has demonstrated that these advances can be used to improve dry scrubber SO_2 removal and sorbent utilization to levels previously attainable only by wet scrubbers. The next stage in the commercialization of these advances is the DOE Combustion 2000 Program discussed below.

LIDS and Combustion 2000

B&W is currently involved in the DOE Combustion 2000 Program with the purpose of developing an emissions control system capable of reducing SO_x emissions to one-third of that allowed under the current New Source Performance Standards (NSPS) and particulate emissions to one-half of the NSPS while addressing the concerns of solid waste generation and air toxics regulation. The work

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is being performed as an integral part of B&W's development of an advanced low-emission boiler system in a project entitled, "Engineering Development of Advanced Coal-Fired Low Emission Boiler Systems (LEBS)." The program is sponsored by the U.S. Department of Energy's Pittsburgh Energy Technology Center. The overall goal of the program is to dramatically improve environmental performance and thermal efficiency of conventional, Rankine cycle, coal-fired power plants.

The LEBS program comprises four distinct phases, with a total duration of 7 years (Fall 1992 - Fall 1999). B&W's first phase was completed in August 1994, and included the assessment of candidate subsystems and technologies, developing a research and development plan for the entire program, and preparing a preliminary design of a commercial generating unit. For the control of SO_x and particulate emissions, numerous near-term advanced flue gas cleanup options were assessed and evaluated. On the basis of these results and other qualitative considerations, the B&W LIDS process was selected for further development and evaluation in B&W's LEBS project.

The second phase consists of engineering development, pilot-scale testing in B&W's 5 MBtu/hr SBS facility, and testing in B&W's new 100 MBtu/hr Clean Environment Development Facility (CEDF). The SBS tests are scheduled for January 1995, and the CEDF tests are scheduled for January 1996. Phase three entails the design of a 10-50 MW_{th} proof-of-concept test facility that will be built and tested in phase four.

Economic Analysis

Results from this project were used to estimate the operating costs of Advanced Dry Scrubbing, LIDS, and A+LIDS, and compare them to more conventional FGD technologies such as wet scrubbing and SCR. The analysis was based on the method described in the U.S. DOE Program Opportunity Notice DE-PSO1-88FE61530, and EPRI's Technical Assessment Guide (TAG)[4].

Plant sizes of 100, 250, and 500 MW, and coal sulfur of 1.5, 2.5, and 3.5% were investigated. The calculations were based on a new plant design with a 30-year book life and 85% capacity factor. Operating costs were also estimated for T_{as} of 10°F and 30°F to illustrate the effect of low T_{as}.

Table 15 shows the annual levelized costs for LIDS, Dry Scrubbing, and Wet FGD operating at **90%** SO_2 removal. Costs are presented in $/ton of SO_2 removed and $/ton coal burned. Figure 24 illustrates the effect of operating at low T_{as} for a 500 MW plant burning 3.5% S coal. At a T_{as} of 30°F, the operating costs for dry scrubbing and LIDS are higher than for wet FGD. However, at a T_{as} of 10°F, operating costs for these technologies are lower than for wet FGD. The cost of wet FGD is $272/ton of SO_2 compared to $243/ton and $259/ton for LIDS and Dry FGD respectively.

Table 16 shows the annual levelized costs for LIDS, Dry Scrubbing, and Wet FGD operating at **95%** SO_2 removal. Costs are presented in $/ton of SO_2 removed and $/ton coal burned. Figure 25 illustrates the effect of operating at low T_{as} for a 500 MW plant burning 3.5% S coal. At a T_{as} of 30°F, the operating costs for LIDS ($303/ton) are much higher than for wet FGD ($258/ton). At a T_{as} of 10°F, operating costs for these technologies are about equal to wet FGD costs with wet FGD at $258/ton of SO_2, $262/ton for LIDS, and $268/ton for Dry FGD. More importantly, this project demonstrated that it is reasonable to even consider the costs of dry FGD systems at SO_2 performance levels and high sorbent utilization previously reserved to wet scrubbers.

TABLE 15 Annual Levelized Costs @ 90% SO$_2$ Removal

$/ton SO$_2$ Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=30°F	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	565	533	597	564	895
100	2.5	443	395	443	403	658
100	3.5	390	327	377	332	494
250	1.5	446	414	437	404	640
250	2.5	371	323	349	309	421
250	3.5	339	276	311	266	327
500	1.5	406	374	383	350	504
500	2.5	347	299	316	276	337
500	3.5	322	259	288	243	272

$/ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=30°F	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	15.3	14.4	16.1	15.2	24.2
100	2.5	19.9	17.8	19.9	18.1	29.6
100	3.5	24.6	20.6	23.8	20.9	31.1
250	1.5	12.0	11.2	11.8	10.9	17.3
250	2.5	16.7	14.5	15.7	13.9	18.9
250	3.5	21.4	17.4	19.6	16.8	20.6
500	1.5	11.0	10.1	10.3	9.5	13.6
500	2.5	15.6	13.5	14.2	12.4	15.2
500	3.5	20.3	16.3	18.1	15.3	17.1

Basis: 90   % SO$_2$ Removal

New Plant Design

30 yr book life, 85% Capacity Factor

FIGURE 24   Effect of T_{as} on Costs for a 500 MW Plant,
3.5 % S Coal and 90% SO_2 Removal
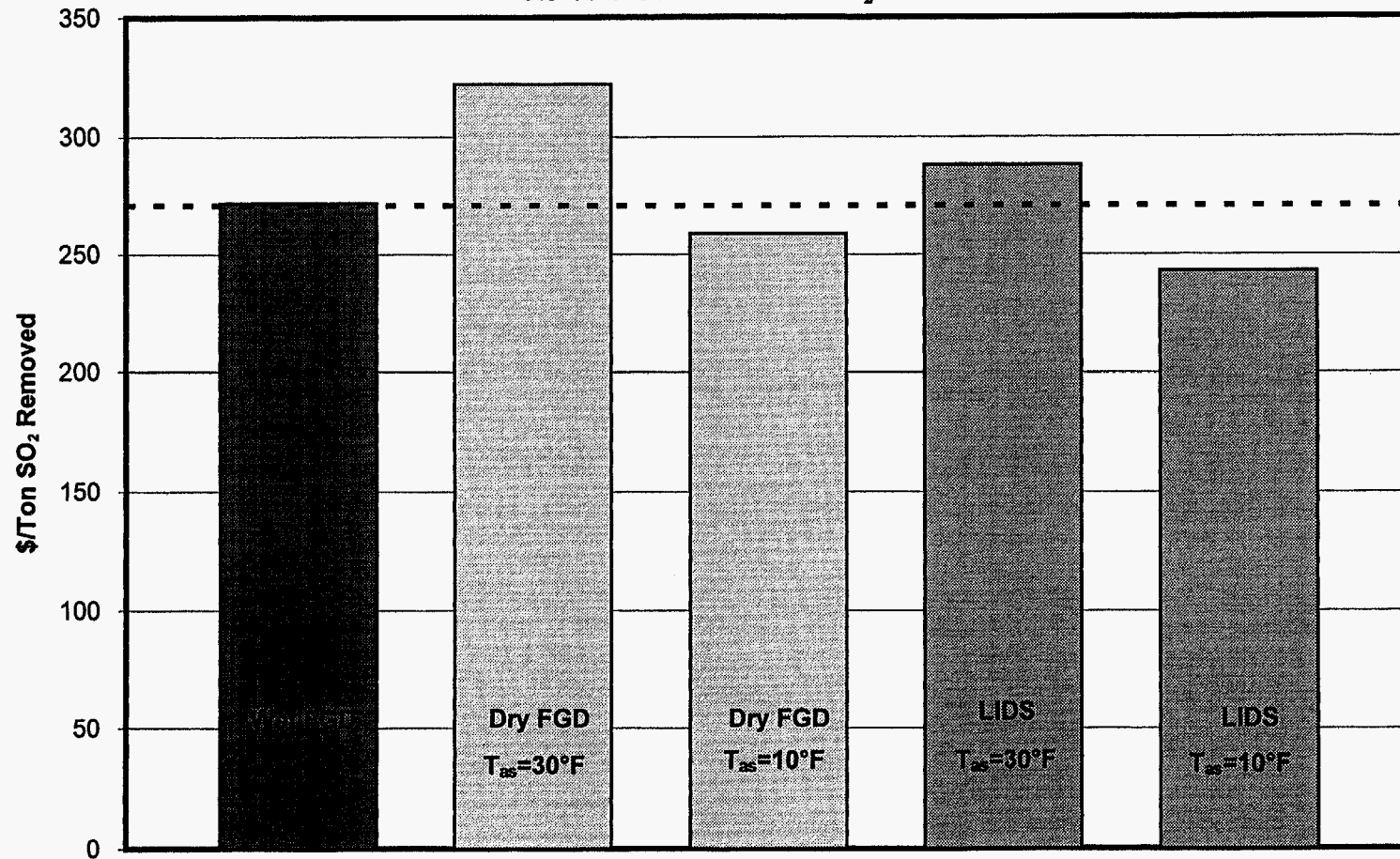
TABLE 16 Annual Levelized Costs @ 95% SO$_2$ Removal

$/ton SO$_2$ Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	527	596	561	855
100	2.5	397	450	411	631
100	3.5	332	388	346	474
250	1.5	415	445	387	616
250	2.5	329	361	322	406
250	3.5	284	325	284	316
500	1.5	377	393	359	486
500	2.5	306	330	291	326
500	3.5	268	303	262	258

30 yr book life, 85% Capacity Factor

$/ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD T_{as}=10°F	LIDS T_{as}=30°F	LIDS T_{as}=10°F	Wet FGD
100	1.5	15.0	17.0	16.0	24.4
100	2.5	18.9	21.4	19.5	30.0
100	3.5	22.1	25.8	23.0	31.5
250	1.5	11.8	12.7	11.0	17.6
250	2.5	15.6	17.1	15.3	19.3
250	3.5	18.9	21.6	18.9	21.0
500	1.5	10.7	11.2	10.2	13.9
500	2.5	14.5	15.7	13.8	15.5
500	3.5	17.8	20.1	17.4	17.2

Basis: 95   % SO$_2$ Removal

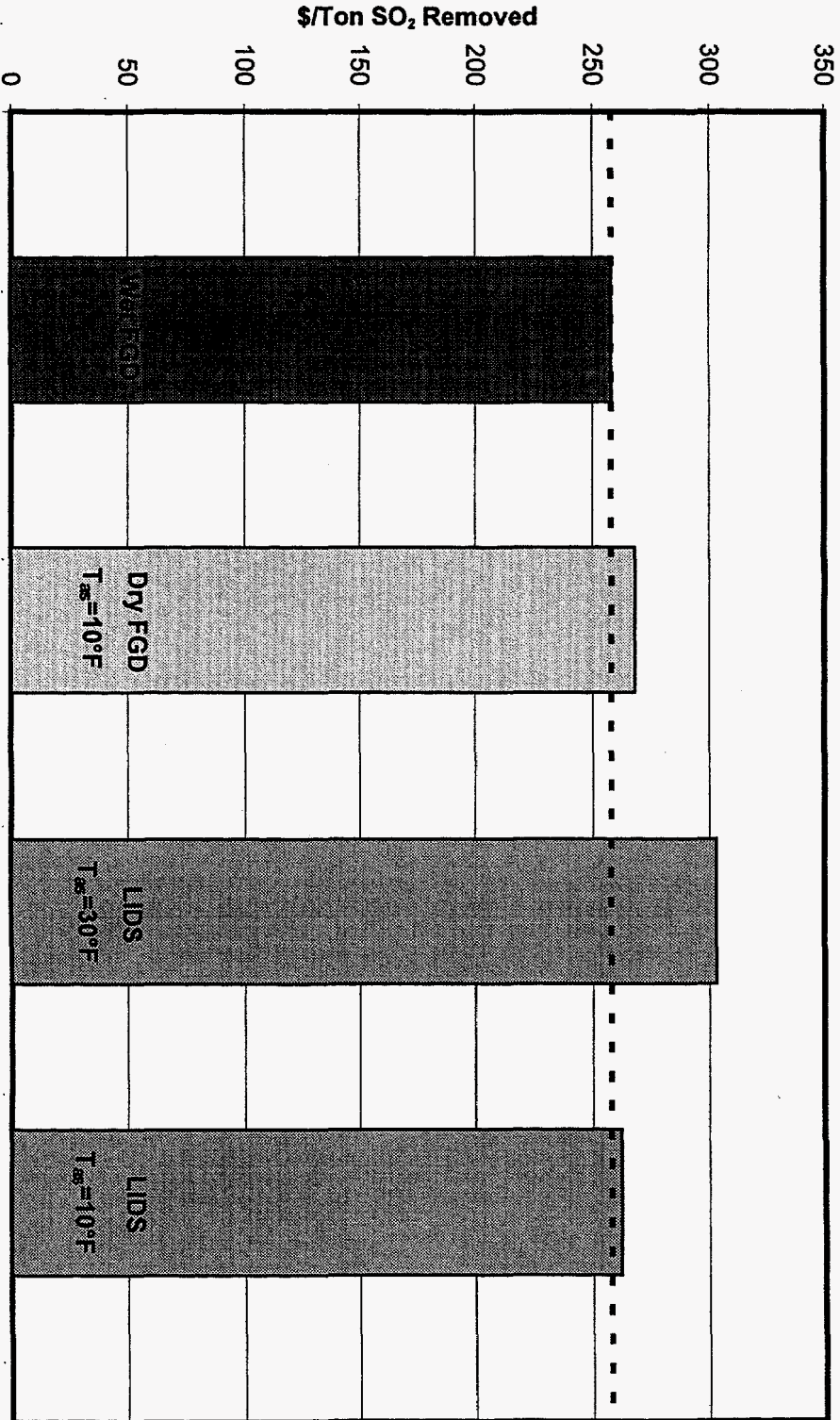New Plant Design

30 yr book life, 85% Capacity Factor

FIGURE 25 Effect of T_{as} on Costs for a 500 MW Plant, 3.5% S Coal and 95% SO_2 Removal

Table 17 shows the annual levelized costs for A+LIDS and a system combining selective catalytic reduction (SCR), Dry FGD, and a fabric filter (FF). Calculations assumed 90% SO_2 and 60% NO_x Removal. Costs are presented in \$/ton of SO_2 removed and \$/ton coal burned. Figure 26 shows that for a 500 MW plant burning 3.5% S coal, A+LIDS costs less than the SCR system at \$287/ton and \$337/ton of SO_2 removed, respectively.

Waste disposal costs for Advanced Dry Scrubbing and LIDS were also estimated, by the method described above, for plants burning coals containing 1.5%, 2.5%, and 3.5% sulfur and scrubbing at 90% efficiency. For dry scrubbing, disposal costs ranged between \$45.50 and \$56.00 per ton of SO_2 removed at a T_{as} of 30°F for low- and high-sulfur coal respectively. By operating at a T_{as} of 10°F, disposal costs could be reduced by 15% and 25% to \$38.50 and \$42.00 per ton of SO_2 removed. Converting to \$/ton of coal burned gives \$1.52 and \$4.36 at a T_{as} of 30°F, and \$1.28 and \$3.27 at a T_{as} of 10°F for the same cases as above. For LIDS (and A+LIDS), disposal costs ranged between \$70.00 and \$87.50 per ton of SO_2 removed at a T_{as} of 30°F for low- and high-sulfur coal respectively. By operating at a T_{as} of 10°F, disposal costs could be reduced by 10% and 20% to \$63.00 and \$70.00 per ton of SO_2 removed. Converting to \$/ton of coal burned gives \$2.10 and \$5.44 at a T_{as} of 30°F, and \$2.33 and \$6.81 at a T_{as} of 10°F for the same cases.

**TABLE 17   Annual Levelized Costs @ 90% SO_2, 60% NO_x Removal**

\$/Ton SO_2 + NO_x Removed

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD, SCR & FF	A+LIDS T_{as}=10°F
100	3.5	405	371
250	3.5	354	309
500	3.5	337	287

\$/Ton of Coal

Plant Size MW$_e$	Coal Sulfur wt. %	Dry FGD, SCR & FF	A+LIDS T_{as}=10°F
100	3.5	26	23
250	3.5	22	19
500	3.5	21	18

Basis: 90   % SO_2 Removal

          60   % NO_x Removal

New Plant Design

30 yr book life, 85% Capacity Factor

FIGURE 26   Dry FGD/SCR/FF and A+LIDS Costs, 500 MW, 3.5% S 90% SO$_2$ and 60% NO$_x$ Removal
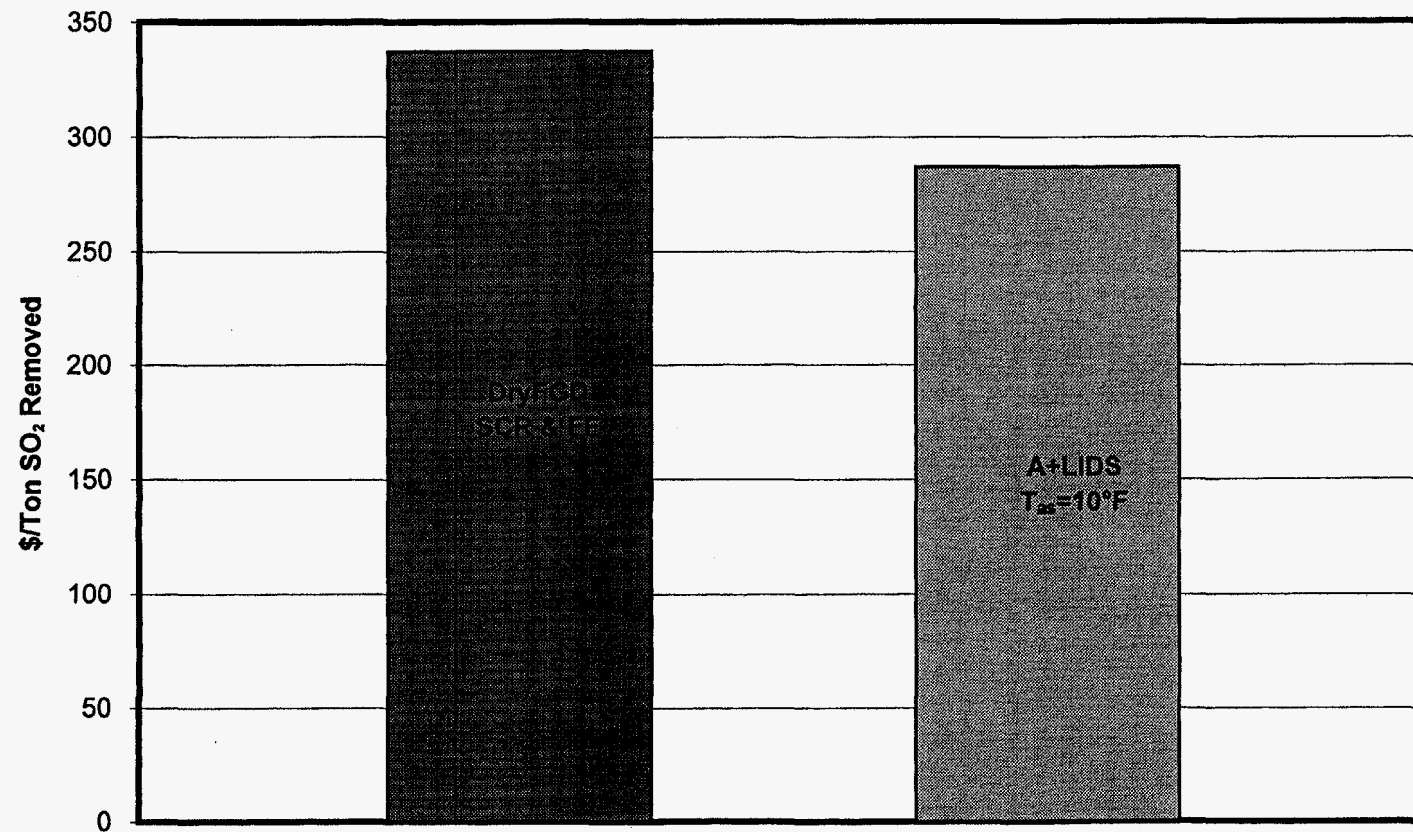
2.4 Final Budget

The final budget is shown in Table 18. Not included in this table is $35,000 contributed by B&W during Phase 5 to investigate the poor performance attained during initial tests. With this money, it was discovered that the addition of small amounts of $CaCl_2$ to the slurry sufficiently slowed evaporation and returned performance to normal levels.

TABLE 18   Project Cost Summary

	Budget	Actual
Personnel	$286,831	$287,802
OCDO	$143,416	$142,677
B&W	$143,415	$145,125
Supplies	$62,957	$69,747
OCDO	$32,198	$35,288
B&W	$30,759	$34,459
Travel	$1,438	$829
OCDO	$0	$0
B&W	$1,438	$829
In-Kind Contribution	$69,000	$69,000
OCDO	$34,500	$34,500
B&W	$34,500	$34,500
Indirect Total	$826,725	$828,873
OCDO	$413,362	$411,011
B&W	$413,363	$417,862
TOTAL COST	$1,246,951	$1,256,251
OCDO	$623,476	$623,476
B&W	$623,475	$632,775

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APPENDIX A
DESCRIPTION OF THE PILOT FACILITY

The proposed Pilot Facility is shown in Figure A-1. Most of the equipment is part of a pilot-scale combustion furnace called the Small Boiler Simulator (SBS). Other equipment, installed during the OCDO sponsored LIDS project, consists of the dry scrubber, detention slaker, and baghouse. The SBS has been used by B&W for coal and burner characterization, gas reburning, furnace sorbent injection, low-NO$_x$ burner tests, and other combustion tests. The main components of the SBS facility consist of the coal-preparation and feeding system, SBS furnace, and gas-monitoring system. New equipment installed for this project include a continuous-slurry preparation system, a loss-in-weight-feeder, and slurry strainers. Each of these components are discussed in detail below.

SBS Furnace - Figure A-2 illustrates the SBS which is rated a 5 million Btu/hr. The SBS is a wall-fired, vertical furnace that simulates the characteristic geometry of B&W's front-wall, coal-fired boilers. At maximum capacity the furnace has a nominal residence time of 1.5 seconds. The quench rate for this unit is 1200°F/s which is similar to that of commercial units.

The inside surface of the furnace is insulated to generate a furnace exit gas temperature of 2250°F. The furnace has five water jacket modules located at the primary furnace, lower furnace, middle furnace, furnace arch, and upper furnace. The primary furnace is insulated with KAO-PHOS93 and the rest of the surfaces are insulated with three inches of Kaowool blanket. Observation ports at several locations are used to monitor the combustion and sorbent injection processes. Primary air is dampered at the FD fan and measured at an orifice ahead of the air heater. The air heater is controlled to maintain a fuel/air mixture temperature between 130°F and 150°F. A gas-fired indirect air heater is used to heat the secondary air to temperatures up to 800°F.

The convection pass consists of four sections. The first contains 42 tubes separated by a large spacing to simulate the secondary superheater. The reheater, primary superheater, and economizer sections contain 40, 100, and 216 tubes, respectively. An atmospheric steam drum provides natural recirculation of cooling water for the tube banks and walls in the convection pass. Soot blowers are located in front of each tube bank. Observation ports along the convection pass can be used as flue gas sampling ports and to observe the deposition on the tube banks.

Heat Exchanger - A heat exchanger at the exit of the SBS cools the flue gas leaving the SBS from about 700°F to 300°F. This is representative of air heater exit temperatures of most commercial units. The heat exchanger is a water-cooled, U-Shape, single pass unit. Water is fed to the jacket counter currently at a rate of 5-10 gpm. The flow rate is controlled to maintain a constant scrubber inlet gas temperature.

Paste Slaker - A Wallace & Tiernan paste slaker is part of a dry scrubbing research facility constructed during the development of B&W's horizontal dry scrubber. The slaker system is rated to 8000 lb/h and is completely equipped with controls, pumps, mixers and storage tanks.

Detention Slaker - A detention slaker was used during LIDS, A+LIDS and Dry MgO Scrubbing research. The detention slaker is a simple, stirred 500 gallon tank. The tank, mixer, and slurry pumping system existed from past projects. The main slurry pump is an existing progressive cavity Moyno pump rated at 16 gpm.
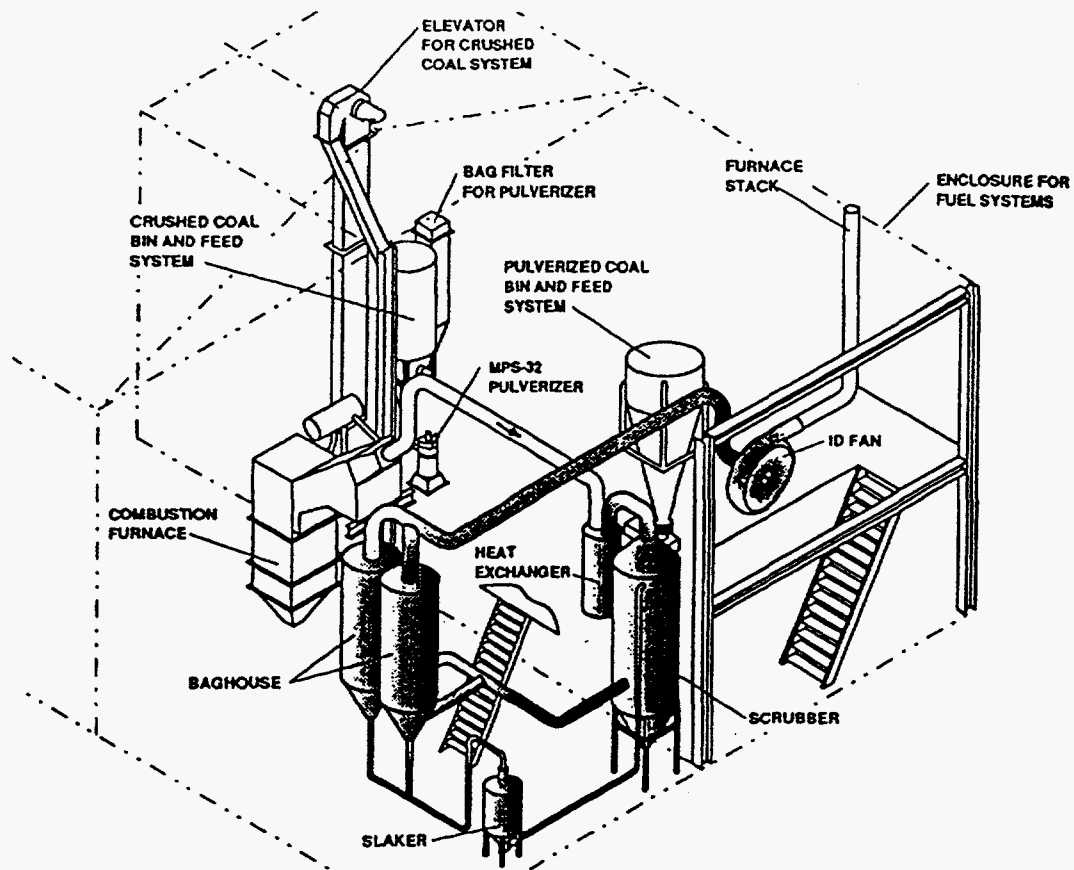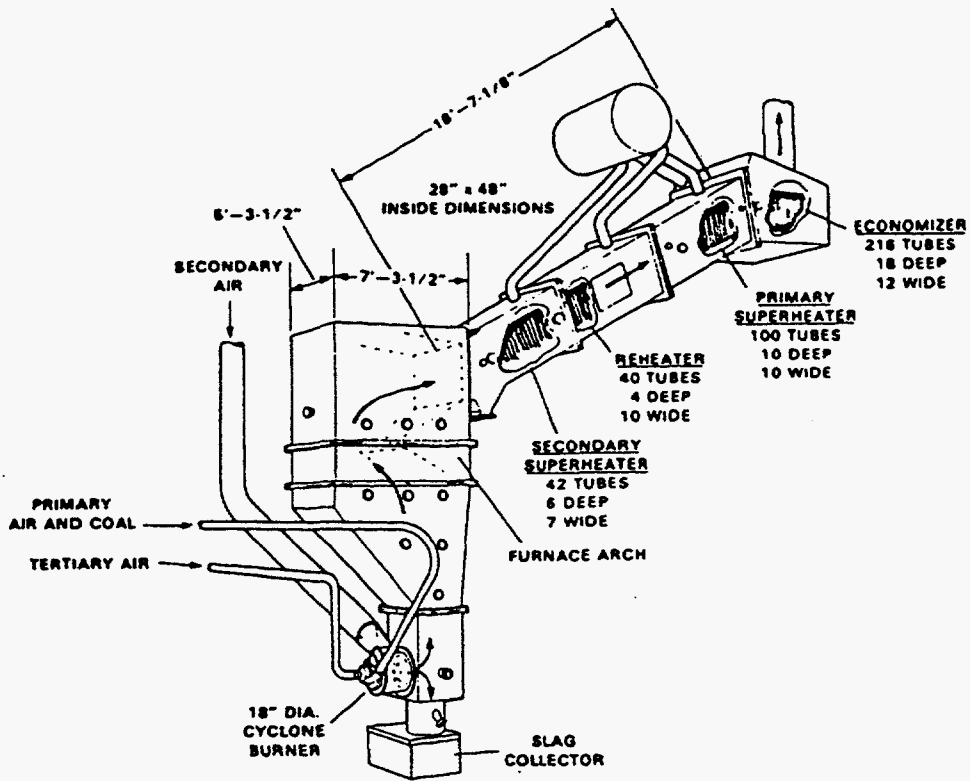
A-1

FIGURE A-1   Pilot Facility

FIGURE A-2   Small Boiler Simulator (SBS)

A-2

Dry Scrubber - The dry scrubber was designed and built as part of the 1990 OCDO Pilot LIDS Demonstration (OCDO Grant CDO/D-87-60). A schematic is shown in Figure A-3. The scrubber is a cylindrical, down-flow reactor measuring 5 feet in diameter and 17 feet long from the atomizer tip to the exit. Flue gas enters the top through an expansion containing several flow straighteners. A single B&W DuraJet^{tm} atomizer, rated at 0.8 gpm, is installed at the top center of the scrubber. Flue gas/slurry mixing is achieved by the rapid entrainment of the high velocity atomized jet stream. Flue gas exits near the bottom of the chamber and flows to the baghouse. The reaction chamber is designed for a gas residence time of 10-12 seconds. The diameter is large enough to prevent direct-wall impingement of the atomized spray. A hopper is located at the base of the scrubber to facilitate solids handling.

The scrubber is completely instrumented with thermocouples, gas analyzers, manometers, and flow meters. Data acquisition and reduction, including energy and water balances, is accomplished by LabTech Notebook software. This software integrates the data received from remote sources by a Fluke Helios Mainframe with an EXCEL spreadsheet. This equipment was purchased and programmed during the OCDO Pilot LIDS Demonstration. Only minor modifications were required for this project. LabTech allows real time observation of important process variables and stores a complete set of all process variable upon command. A similar DAS system is dedicated to the SBS to monitor, control, and store pertinent furnace data.

Baghouse - The MikroPul baghouse is shown in Figure A-4. It consists of two modules each containing 23, 4-5/8 inch diameter, 10 feet long bags. Several new bag materials were tested during this project to determine which type of bag has the best cleaning characteristics at low T_{as}. The modules are insulated to allow the SBS to limit heat loss. A 25 HP, 1500 SCFM, 24 in. w.g., ID balances the pressure drop created by the flues, heat exchanger, dry scrubber, and baghouse. An inlet damper is automatically controlled to balance the draft created by the FD fan of the SBS facility.

Gas Monitoring System - The gas analysis system installed on the SBS facility consists of Beckman Analyzers. These analyzers continuously monitor and record the stack concentrations of O_2, SO_2, NO_x, CO, and CO_2. The system is equipped with calibration gas that can be injected into the stack or at the inlet of the analyzers. Particulate is filtered from the flue gas at the point of sampling. The gas sample is dehumidified by a refrigerated condenser and by a Purma Pure dryer prior to the analyzers. This system monitors gas composition at the exit of the SBS prior to the dry scrubber.

A separate gas sampling system was used to monitor the SO_2 and O_2 concentrations at the exit of the dry scrubber and baghouse. This system consists of Anarad SO_2 and Teledyne O_2 analyzers. The gas extraction probe is similar to the SBS system in that particulate is removed at the point of sampling and the water condensed before the gas enters the analyzers. B&W has an extensive background in the installation and use of this type of system from previous developmental work of dry scrubbers. During Phase 6 - A+LIDS testing, a Severn Science ammonia analyzer was used to monitor the ammonia concentration in the flue gas at the exit of the baghouse.

Facility Upgrade - As part of this project, the facility was upgraded to allow continuous operation of the slurry preparation and recycle system. The system, shown in Figure A-5, consists mainly of two baghouse collection hoppers, an automatic batch weighing and ash transport system, two detention slaking tanks, and a slurry storage tank. Material from the baghouse can be transported
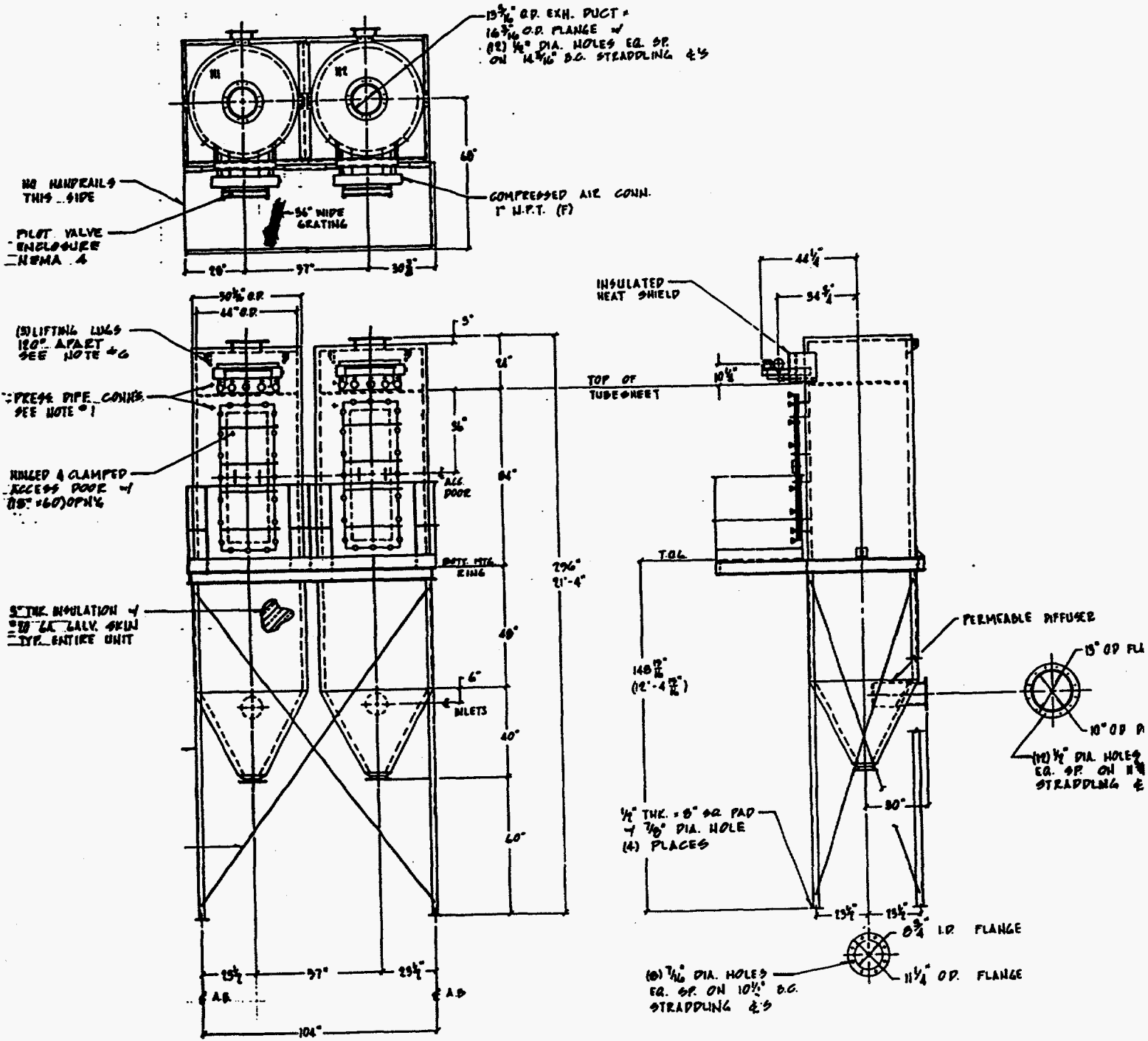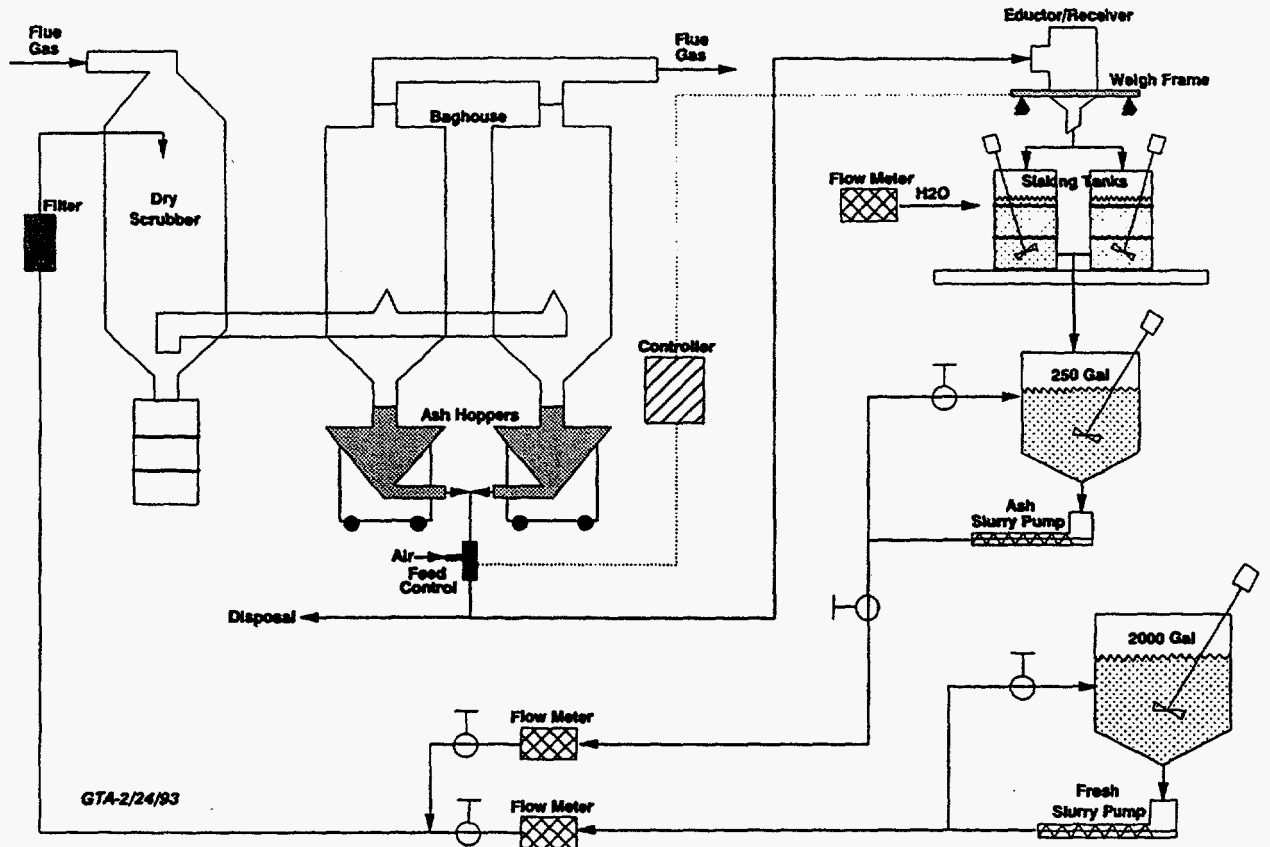
FIGURE A-3   Schematic of the Dry Scrubber

A-4

FIGURE A-4   Schematic of the Baghouse

SBS Dry Scrubber Continuous Slurry System

to the slakers or a waste hopper via a diverter valve. Two slaking tanks are required because detention slaking requires a minimum of one hour. One tank continuously slakes while the other is being filled. The 500 gallon storage tank has sufficient capacity to operate continuously for several hours.

Other modifications included the installation of a wet strainer between the slaking tanks and the storage tank, and the purchase of a new loss-in-weight feeder. The strainer removes large pieces of dried slurry that periodically form and break off from the sides of the slaking tanks. Large pieces can plug the atomizer and seriously effect atomization quality if not removed. The loss-in-weight feeder meters the limestone injected into the furnace. The new feeder greatly increases the accuracy and reproducibility of this important test variable.