High SO₂ Removal Efficiency Testing

Final Report

July 24, 1992 through October 31, 1997

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October 15, 1997

Project Number DE-AC22-92PC91338
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SUMMARY

This final report describes the results of performance tests at six full-scale wet lime- and limestone-reagent flue gas desulfurization (FGD) systems. The objective of these tests was to evaluate the effectiveness of low capital cost sulfur dioxide (SO₂) removal upgrades for existing FGD systems as an option for complying with the provisions of the Clean Air Act Amendments of 1990. The upgrade options tested at the limestone-reagent systems included the use of organic acid additives (dibasic acid (DBA) and/or sodium formate) as well as increased reagent ratio (higher excess limestone levels in the recirculating slurry solids) and absorber liquid-to-gas ratio. One system also tested operating at higher flue gas velocities to allow the existing FGD system to treat flue gas from an adjacent, unscrubbed unit. Upgrade options for the one lime-based system tested included increased absorber venturi pressure drop and increased sulfite concentration in the recirculating slurry liquor.

Table S-1 summarizes the SO₂ removal performance data from the six test sites. In this table, the “baseline” SO₂ removal efficiency for each site is the measured efficiency of the individual module tested when operating at normal full-load conditions. This efficiency may not be the same as the overall FGD system efficiency because several of the sites use partial flue gas bypass to control outlet emissions at the permitted level. The “maximum” SO₂ removal efficiency in the table is the maximum that was measured during the upgrade tests, generally conducted on the same module. This value may be slightly lower than the absolute maximum efficiency that could be achieved with that absorber module, because the test conditions were not extended to achieve a true maximum in most cases. Also, this maximum does not consider the effects of any flue gas bypass. The "economic" optimum removal efficiency is the efficiency above which additional SO₂ removal was not estimated to be cost effective. This efficiency was based on what could be obtained using the most cost-effective upgrade option for that site, and a maximum incremental cost for additional SO₂ removed of $150/ton.

The results in Table S-1 show that the target removal efficiency of 95 to 98% for this project was achieved at five of the six sites. For sites that currently bypass a portion of the flue gas, achieving this target removal efficiency would require that the bypass be either closed or
Table S-1
Summary of SO$_2$ Removal Performance and Costa Data for the Six Test Sites

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Absorber Type</th>
<th>Reagent Type</th>
<th>Oxidation Mode</th>
<th>Upgrade Tested</th>
<th>Baseline % SO$_2$ Removal</th>
<th>Maximum % SO$_2$ Removal$^a$</th>
<th>Economic Optimum % SO$_2$ Removal$^b$</th>
<th>Average Cost of Add'l SO$_2$ Removal ($/ton)</th>
<th>Additional SO$_2$ Removed (tons/yr)</th>
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<td>Limestone</td>
<td>Forced</td>
<td>DBA</td>
<td>93</td>
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<td>99</td>
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<td>380,000</td>
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<td>Limestone</td>
<td>Inhibited</td>
<td>DBA Sodium Formate</td>
<td>86</td>
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<td>(94) c</td>
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<td>97</td>
<td>76</td>
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<td>790,000</td>
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</table>

(a) This is the maximum SO$_2$ removal efficiency observed during tests. It is not the maximum possible SO$_2$ removal.

(b) The economic SO$_2$ removal is defined as the removal efficiency above which the cost or additional removal exceeds $150/ton of SO$_2$.

(c) Cost of additional SO$_2$ removal is negative (a savings is obtained) because of increased reagent utilization.
reduced to a minimal value. The only site that did not achieve the target removal efficiency was at Duquesne Light’s Elrama FGD system, for which the maximum removal efficiency was about 93%. However, the venturi absorber type at this site had a relatively limited gas/slurry contact area that in turn limited SO₂ removal, and there was no low-capital cost upgrade that could improve on that basic limitation. The maximum SO₂ removal efficiency exceeded 99% at three of the six sites.

For the three FGD systems other than Elrama that had baseline removal efficiencies in the 85 to 90% range, the economic optimum SO₂ removal efficiencies ranged from 95 to 97%. For the two systems that had baseline efficiencies higher than 90%, the optimum removal efficiency was 99%.

Table S-1 also shows the estimated quantity and average cost of additional SO₂ that could be removed using the most cost-effective upgrade option at each site. The "average cost of additional SO₂ removal" is the total annual increase in operating costs (above baseline costs) when using the upgrade option, divided by the total tons of additional SO₂ removed (above baseline removal). In most cases, these additional operating costs included increased reagent use, increased byproduct solids disposal costs, additive purchase costs, and the annualized cost of new additive storage and handling equipment. For the Tampa Electric site, where wallboard-grade gypsum is produced, additional revenue is actually derived from increased SO₂ removal and the resulting gypsum solids production. This revenue was considered in calculating the costs for this site. At the Duquesne Electric site, the most cost-effective upgrade option actually improves lime utilization, so a savings in reagent cost would be realized. For all of the sites, other minor differences in operating costs due to changes in flue gas pressure drop or in recycle pump power consumption were included where applicable.

The results in Table S-1 show that increasing the efficiency of these existing FGD systems should be very cost effective if SO₂ allowances are valued at $150/ton. The average cost for additional SO₂ removal for five of the six sites ranged from $39 to $76/ton. A net operating cost savings could be obtained at the Elrama FGD system due to increased reagent utilization. The annual value of allowances that could be generated using the upgrade options that were tested in the project were estimated to range from about $400,000 to $2.4 million (assuming a $150/ton value).
To date, two of the six utilities have used the results of this project as a basis for implementing upgrades to their existing FGD systems. It is likely that the other four will use these results as they formulate a strategy for achieving compliance with Phase II of the 1990 Clean Air Act Amendments, in the year 2000.
1.0 INTRODUCTION

1.1 Background

Provisions of the 1990 Clean Air Act Amendments call for a ten-million ton per year reduction in U.S. sulfur dioxide (SO₂) emissions (from a 1980 baseline) in two phases. Phase I calls for a five-million ton per year reduction that was implemented by 1995, and the remainder of the reductions are to be completed by the year 2000 for Phase II. Affected utilities have a number of options for achieving these reductions, such as switching to lower sulfur-content coals, purchasing SO₂ allowances from other utilities, installing new flue gas desulfurization (FGD) systems, and improving the SO₂ removal performance of existing FGD systems. Some utilities may employ a combination of these and other options as part of an overall compliance strategy.

The Flue Gas Cleanup Program at the U.S. Department of Energy Federal Energy Technology Center (DOE FETC) helps to maintain and foster the widespread use of coal by developing technologies that will mitigate the environmental impacts of coal utilization. The program focuses on post-combustion technologies for the control of SO₂, oxides of nitrogen, fine particulate matter, and air toxics generated from coal combustion. A portion of the program, including this project, involves enhancing the SO₂ removal capabilities of existing wet FGD systems. The results from this project will allow utilities to better consider enhanced performance of existing FGD systems as an option for achieving compliance with Phase II of the 1990 Clean Air Act amendments.

In this project, Radian International LLC conducted tests at six full-scale FGD systems to evaluate options for achieving high SO₂ removal efficiencies (the target was 95 to 98% removal). Each system was characterized under baseline operation, and then with low capital cost modifications to enhance SO₂ removal performance.
The systems evaluated were at the Tampa Electric Company's (TECo's) Big Bend Station, Hoosier Energy's Merom Station, the Southwestern Electric Power Company's (SWEPCo's) Pirkey Station, PSI Energy's Gibson Station, Duquesne Light's Elrama Station, and the New York State Electric and Gas Corporation's (NYSEG's) Kintigh Station. A wide variety of FGD system vendors and designs were represented in the project. Most of these systems were originally designed to achieve 85 to 90% SO₂ removal.

This final report provides an overview of results from each of the six test sites. Overall performance results and SO₂ removal upgrade economics are emphasized. Separate topical reports have been issued for each of the individual sites. The reader is referred to these topical reports for more detailed descriptions of the test results and for more detailed test data summaries.

1.2 Project Description

Three types of performance tests were completed at each of the sites. First, baseline tests were conducted to obtain performance data without the SO₂ removal upgrades. Then, parametric tests were done to obtain performance data using various upgrade options over a range of operating conditions. The baseline and parametric tests were conducted using only one of the several operating gas absorber modules in each of the FGD systems. Following the parametric tests, an extended steady-state test was conducted at most sites to demonstrate performance of the most cost-effective upgrade when applied to the entire FGD system. At the systems where organic acid additives were used, additive consumption was measured during the extended tests.

Under a separate project funded by the Electric Power Research Institute, their FGD Process Integration and Simulation Model (FGDPRISM) was calibrated to each of the six FGD systems using the test data. The calibrated models were then used to predict system SO₂ removal performance over a range of process conditions, including conditions that could not be tested.

Economic calculations were conducted to determine the most cost-effective approach for achieving the project target of 95 to 98% SO₂ removal. Actual and predicted
performance results, along with measured steady-state additive consumption data, additive system capital cost estimates, and other pertinent cost information provided by the participating utilities, were the basis for the economic evaluations. In these evaluations, the net cost for additional tons of SO₂ removal was estimated for different operating conditions. These costs can be compared with the expected market value of SO₂ allowances, or the expected cost of allowances generated by other means such as fuel switching or new scrubbers, to arrive at the most cost-effective strategy for compliance with the Clean Air Act amendments.

1.3 Report Organization

The remainder of this report includes four sections. Section 2 provides a basic overview of wet FGD technology. Section 3 provides a summary of the test results from each of the six FGD systems. FGDPISM modeling results are described in Section 4, and the economic evaluation results are described in Section 5. Section 6 provides a list of references for topical reports previously written for this project.
2.0 DESCRIPTION OF THE FGD PROCESS

This section provides a brief description of the wet FGD process as it is applied in the U.S electric utility industry. This section will be helpful to any reader who does not have at least a cursory understanding of wet FGD technology.

Figure 2-1 provides a simplified flow diagram for the process. Wet FGD systems treat flue gas from coal or other sulfur-containing fossil fuel combustion, and are usually installed downstream of the boiler air heater and particulate control device. Therefore, the gas treated is typically at 250°F to 350°F, near atmospheric pressure, and contains only a small amount of particulate matter (0.1 lb/10^6 Btu or less).

This flue gas enters an absorber vessel (or one of several vessels installed in parallel) where it is contacted with a recirculating aqueous liquor or, more commonly, aqueous slurry. All of the systems tested in this program use slurry processes, so the remaining discussions will focus on slurries only.

As the flue gas contacts the slurry, a portion of the water in the slurry cools and saturates the gas, and SO₂ is absorbed from the flue gas into the slurry liquor. A number of techniques are used to provide intimate gas/slurry contact in the absorber vessel, such as finely atomized slurry sprays, packing, and/or contactor trays. The slurry is usually introduced near the top of the gas contacting section of the absorber and flows downward due to gravity. It can be recirculated as part of a single loop, or in multiple, separate slurry loops. The gas flow through this section can be upward (countercurrent to the slurry), downward (co-current with the slurry) or horizontal (cross-flow relative to the slurry) depending on the system design. After the scrubbed gas leaves the contact section, it flows through a mist eliminator to remove entrained slurry droplets, then usually out through a chimney or stack to the atmosphere.

The slurry falls to the bottom of the absorber after contacting the flue gas and returns to a reaction tank. In the reaction tank, an alkaline reagent is added to the slurry to enhance its ability to remove SO₂ before it is returned to the absorber. The alkaline reagent is
Figure 1. Simplified Flow Diagram for a Lime/Limestone Reagent FGD System
typically either slaked lime (Ca(OH)$_2$) or finely ground limestone (CaCO$_3$) and is also added to the reaction tank as a slurry. The makeup rate of this reagent slurry is controlled to maintain the pH of the recirculating slurry in the reaction tank at a set value.

Some systems use additives to increase the liquid-phase alkalinity in the recirculating slurry feed to the absorber. These additives enhance SO$_2$ removal performance because the lime or limestone reagent solids in the slurry may not dissolve quickly enough to provide sufficient alkalinity in the absorber. Additives include magnesium (usually added as magnesium oxide) and certain organic acids such as formic acid (usually added as sodium formate) or "dibasic acid" (DBA). The latter is a mixture of adipic, glutaric, and succinic acids. Magnesium increases liquid-phase alkalinity by allowing higher sulfite ion concentrations in the liquor; sulfite is an alkaline species with respect to SO$_2$ removal. Organic acid additives enhance liquid-phase alkalinity by serving as a buffer.

The byproduct of SO$_2$ removal in the absorber is either calcium sulfite hemihydrate (CaSO$_3$·½H$_2$O) in inhibited- or low-oxidation systems or calcium sulfate dihydrate (CaSO$_4$·2H$_2$O, also called gypsum) in forced-oxidation systems. In a "natural" oxidation system, a mixture of these byproducts is typically formed because a portion of the sulfite is oxidized to the sulfate form by oxygen in the flue gas. In inhibited-oxidation systems, oxidation is controlled at 15% or less by adding additives (typically sulfur) to avoid gypsum scaling. In forced-oxidation systems, oxidation is controlled at greater than 90% to produce a high-quality gypsum byproduct by introducing compressed air in the reaction tank.

As SO$_2$ is removed and byproduct solids are generated, the solids content in the recirculating slurry increases, due to water evaporation into the flue gas, byproduct solids formation, and continual reagent makeup. To maintain consistent slurry levels and slurry solids content in the reaction tank, most systems operate with a nearly continuous blowdown of a portion of the recirculating slurry, and with some water makeup. The pH of the recirculating slurry is maintained at a level high enough to achieve the desired SO$_2$ removal performance, but low enough that there is only a small amount of excess reagent lost in the blowdown.
This blowdown slurry is piped to a dewatering system consisting of thickeners, centrifuges, and/or vacuum filters. The water recovered is returned to the absorbers, and the solids recovered are either sent to disposal or sold as a byproduct. Systems that operate at low oxidation percentages generally add fly ash and sometimes lime to stabilize the calcium sulfite byproduct, then landfill the stabilized material. Systems that use forced oxidation to produce gypsum may be able to sell their byproduct gypsum as a feedstock for wallboard manufacturing.

There are numerous potential variations in the basic process described above and illustrated in Figure 2-1. Some of these variations are seen in the diversity of the six systems that were investigated during this project, as described in the next section.
3.0 TEST RESULTS

The primary performance measurement made during tests at the various FGD systems was SO2 removal efficiency. Efficiency was determined by simultaneous measurement of inlet and outlet SO2 concentrations in flue gas treated by a single absorber module. At most of the sites, organic acid additives (DBA or sodium formate) were tested as potential upgrade options. The effects of other operating variables such as slurry pH set point, flue gas velocity, and liquid-to-gas ratio (L/G) on SO2 removal performance were also measured. The SO2 removal performance results are discussed below in Section 3.1.

The cost-effectiveness of additives used to upgrade SO2 removal efficiency depends not only on the concentration required in the FGD recirculating slurry liquor, but also on the corresponding additive consumption rate. At each of the sites where DBA or sodium formate was tested, an extended steady-state test was conducted with the additive being used in the entire FGD system, and the consumption rate was calculated by material balance. Results of these additive consumption tests are summarized in Section 3.2.

Organic acid additives may affect other process operating parameters besides SO2 removal efficiency. The observed effects of additives on reagent utilization, sulfite oxidation, and solids dewatering properties are presented in Section 3.3.

3.1 SO2 Removal Efficiency

Table 3-1 provides a summary of the configurations of the six FGD systems that were included in this project. The table also includes the baseline and the highest SO2 removal performance measured for each system.

Figures 3-1 through 3-7 show typical performance data from each of the six test sites. In these figures, SO2 removal performance is expressed on the Y axes as the "number of transfer units" or NTU, which is a widely used method for expressing gas-liquid mass transfer data. NTU was calculated as ln(SO2in/SO2out) using the inlet and outlet SO2 concentrations measured
Table 3-1
Summary of SO2 Removal Performance Data for the Six Test Sites

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Absorber Type</th>
<th>Reagent Type</th>
<th>Oxidation Mode</th>
<th>Upgrade Tested</th>
<th>Baseline % SO2 Removal</th>
<th>Maximum % SO2 Removal</th>
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<tbody>
<tr>
<td>Tampa Electric</td>
<td>Dual-Loop Spray/Packed</td>
<td>Limestone</td>
<td>Forced</td>
<td>DBA</td>
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<td>Limestone</td>
<td>Inhibited</td>
<td>DBA</td>
<td>86</td>
<td>not tested</td>
</tr>
<tr>
<td>Gibson 5</td>
<td></td>
<td></td>
<td></td>
<td>Sodium Formate</td>
<td>86</td>
<td>97.8</td>
</tr>
<tr>
<td>Duquesne Light</td>
<td>Venturi</td>
<td>Mg-Lime</td>
<td>Inhibited</td>
<td>Venturi DP</td>
<td>89</td>
<td>93.2</td>
</tr>
<tr>
<td>Elrama</td>
<td></td>
<td></td>
<td></td>
<td>Thiosulfate</td>
<td></td>
<td></td>
</tr>
<tr>
<td>New York State Electric and Gas Corp.</td>
<td>Spray</td>
<td>Limestone</td>
<td>Inhibited</td>
<td>Sodium Formate</td>
<td>86</td>
<td>99.4</td>
</tr>
<tr>
<td>Kintigh</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

(a) This is the maximum SO2 removal efficiency observed during tests. It is not the maximum possible SO2 removal.
Figure 3-1. NTU vs. Upper-Loop DBA Concentration at Tampa Electric Big Bend 4

Figure 3-2. Effect of Gas Velocity on SO₂ Removal at Tampa Electric Big Bend 4
Figure 3-3. NTU vs. Additive Concentration at Hoosier Energy Merom (Predicted by FGDPRISM)

Figure 3-4. NTU vs. Upper-Loop Additive Concentration at SWEPCO Pirkey
3 Modules, pH 5.3 - 3 Modules, pH 5.7

4 Modules, pH 5.3 - 4 Modules, pH 5.7

---

98% Removal
L/G = 96 gal/macf

95% Removal
L/G = 96 gal/macf

90% Removal
L/G = 71 gal/macf

80% Removal
L/G = 71 gal/macf

Formate Ion Concentration (ppm)

Absorber NTU

SO₂ removal predicted by FGDPRISM for Baseline Tests 1 and 2 under high inlet SO₂ conditions.

Figure 3-5. NTU vs. Additive Concentration at PSI Energy Gibson 5

---

95.0% Removal
10 in. H₂O

91.8% Removal
12 in. H₂O

86.5% Removal

77.7% Removal

Sulfite Concentration (mg/L)

Absorber NTU

Figure 3-6. NTU vs. Sulfite Concentration at Duquesne Light Elrama
Figure 3-7. NTU vs. Formate Concentration at NYSEG Kintigh
during the tests. For convenience, corresponding SO$_2$ removal efficiencies are shown as dashed lines on the figures. The results for each site are briefly summarized below.

3.1.1 Tampa Electric Company’s (TECo’s) Big Bend Station Unit 4

The TECo FGD system is a forced-oxidation limestone process that employs countercurrent gas flow, dual-loop absorbers. Different operating conditions are maintained in the two (upper and lower) loops. The lower loop has spray headers to provide gas/slurry contact area, while the upper loop has spray headers as well as a layer of open-grid packing. The latter is to improve gas/slurry contacting, and thus enhance SO$_2$ removal performance. Relatively little SO$_2$ removal occurs in the lower loop, which is operated at low pH to maximize limestone utilization in the slurry blowdown from the absorber.

DBA additive was used to increase SO$_2$ removal efficiency at Big Bend. Figure 3-1 shows overall SO$_2$ removal efficiency plotted as a function of upper-loop DBA concentration. The results for the lower pH set point (5.8) show that the overall SO$_2$ removal efficiency could be increased from the baseline level (without additive) of about 93% to nearly 99% with a DBA concentration of about 400 mg/L in the upper loop of the absorber module. At the higher pH set point (6.1), the baseline efficiency was about 96%, and the highest measured efficiency was 99.7% at a DBA concentration of 900 mg/L. Even though the mass transfer efficiency (NTU) continued to increase with increasing DBA concentration throughout the range tested, the marginal increase in SO$_2$ removal percentage was small at high DBA concentrations.

A portion of the baseline and parametric tests at this site were also conducted at higher than normal flue gas velocity through the test absorber. These results are not presented in Figure 3-1, but in general they showed that with DBA additive, the test absorber could achieve high SO$_2$ removal efficiency (up to 99%) when operated at flue gas velocities 33% above the design value (about 10 ft/s vs. 7.5 ft/s), even though the L/G was correspondingly lower. Based on these and subsequent tests conducted by the utility, TECo modified the 485-MW Unit 4 FGD system to also treat flue gas from the adjacent 445 MW Unit 3.
Starting in 1995, TECo now adds 1000 mg/L of DBA to the upper loop and operates all four absorbers (no spare) at flue gas velocities about 50% greater than the design value (nominally 11 ft/s vs. 7.5 ft/s) to treat flue gas from both Unit 4 and Unit 3 in the Unit 4 FGD system. In 1997, this project returned to Big Bend to determine if the Unit 4 FGD system could be successfully operated at even higher velocities, to scrub flue gas from a third unit at the station. To scrub all of the flue gas from three units, the four absorbers would have to operate at superficial gas velocities of up to 17 ft/s, which would be over twice their design velocity.

Prior to conducting the 1997 tests, TECo implemented minor modifications to one absorber to allow it to operate at "ultra-high" velocities. The testing, conducted in June 1997, showed that with the current mist eliminators, the maximum velocity that could be sustained through the absorbers was 13.5 ft/s. The limiting factor was droplet carryover from the mist eliminators. Newer, state-of-the-art mist eliminator designs should allow operation at higher velocities, though.

Parametric testing conducted at the ultra-high velocity conditions showed that high SO₂ removal efficiencies could be maintained. In fact, SO₂ removal was observed to improve with increasing velocity, apparently due to improvements in the overall mass transfer coefficient with velocity. Figure 3-2 illustrates this trend. At most conditions tested during the 1997 parametric tests, the oxidation percentage and limestone utilization values remained high enough to continue producing wallboard-grade gypsum in the test absorber at ultra-high velocities.

An attempt was made to operate the test absorber at ultra-high velocity conditions for an extended period of time. However, during a period of operation with higher sulfur coal, the available oxidation air rate was not sufficient to completely oxidize the removed SO₂, and wallboard-quality gypsum production could not be maintained. The longer-term test at ultra-high velocity conditions was stopped at that point. An upgrade of the oxidation air system capacity would be required to allow continued operation at ultra-high velocity while firing higher sulfur coals.
3.1.2 Hoosier Energy Merom Station Unit 2

The Merom FGD system is an inhibited-oxidation limestone process with cocurrent-gas-flow absorber modules that use open-grid packing. The packing is wetted by slurry distributed through a grid of open pipes above the packing. Both DBA and sodium formate additives were tested at Merom.

Figure 3-3 compares typical performance results for DBA and sodium formate additive at Merom adjusted using a calibrated version of EPRI's FGDPRISM computer model to the same process operating conditions (i.e., the amount of excess limestone reagent in the recirculating slurry, gas velocity, inlet SO₂ concentration). The results show that the maximum SO₂ removal efficiency obtained with the current absorber configuration at Merom is about 97%, even with very high additive concentrations. At high additive concentrations, the performance is "gas-film limited." That is, the SO₂ removal efficiency becomes limited by the amount of packing surface area and by the rate of diffusion of SO₂ from the gas to the surface of the liquid on the packing and therefore unaffected by the liquid composition. Further performance improvements would require increased mass transfer surface area.

3.1.3 Southwestern Electric Power Company's (SWEPCo) Pirkey Station

The SWEPCO FGD system is an inhibited-oxidation limestone process with countercurrent gas flow dual-loop absorbers. The lower loop has slurry sprays to provide gas/slurry contacting, and the upper loop has slurry sprays plus a perforated tray to enhance SO₂ removal. At Pirkey, the additive concentrations in the lower and upper loops are similar because of intentional slurry circulation between the loops. However, the upper and lower slurry pH levels are independently controlled. Significant SO₂ removal occurs in both the lower and upper loops of the absorbers. Both DBA and sodium formate were tested at Pirkey.

Figure 3-4 summarizes SO₂ removal efficiency results at Pirkey for both the DBA and sodium formate additives. Two pH ranges were tested. At the lower pH set point of 5.7, the baseline efficiency was about 86%, and a maximum removal efficiency of about 97% was observed.
with either additive. At the higher pH set point of 6.2, the baseline efficiency was about 97%, and the efficiency could be increased to 99.5% with DBA additive or 99% with sodium formate additive.

3.1.4 PSI Energy’s Gibson Station Unit 5

The Gibson Unit 5 FGD system is an inhibited-oxidation limestone process to which a small amount of dolomitic (i.e., high magnesium content) lime is added to increase the liquid-phase alkalinity and thus enhance SO₂ removal performance. The absorbers use the Kellogg/Weir horizontal cross-flow configuration. Slurry is sprayed into the flue gas through headers and nozzles located along the top of horizontal-gas-flow absorber modules. Each of the four spray headers is supplied by an independent recycle pump, so the liquid-to-gas (L/G) ratio can be controlled at several discrete levels. Sodium formate additive as well as increased reagent ratio (higher excess reagent levels in the recirculating slurry) and increased L/G were tested at Gibson to increase SO₂ removal.

Figure 3-5 summarizes performance results from Gibson. Absorber NTU is plotted as a function of sodium formate additive concentration. Results are shown for two different levels of L/G and two different operating pH levels at each L/G level. At the lower L/G (71 gal/kacf) and lower pH set point (5.3), sodium formate additive increased the SO₂ removal efficiency from a baseline value of about 68% to 86%. At the lower L/G and higher pH set point (5.7), the efficiency increased from a baseline value of about 80% to a maximum value of about 86%.

Results for tests at increased L/G obtained by operating an additional recycle pump showed that NTU increased in proportion to L/G at either operating pH set point. That is, increasing the L/G from 71 to 96 gal/kacf, a 35% increase, correspondingly increased NTU by 35% across the range of formate concentration tested. Tests using different combinations of gas flow and slurry flow to obtain the same L/G (i.e., operating with three modules and four pumps per module versus four modules and three pumps per module) showed that equivalent SO₂ removal efficiency was obtained. At the higher level of L/G, a maximum SO₂ removal efficiency of about 97% was obtained. At any of the four combinations of L/G and slurry pH tested, the formate
additive provided little additional performance enhancement at concentrations higher than about 1500 mg/L.

3.1.5 Duquesne Light Company’s Elrama Station

The Elrama FGD system is a magnesium-enhanced lime process with cocurrent-gas-flow venturi absorbers. In this absorber type, surface area for mass transfer is generated when the recirculating slurry is atomized by the flue gas as the gas flows at high velocity through the venturi absorber throat. The venturi throat cross section is adjustable, providing a range of throat velocities and therefore mass transfer surface area.

Preliminary simulations using EPRI's FGDPRI SM computer model showed that the use of organic acid additives to enhance SO₂ removal would not be cost effective with this process configuration. This was largely because the system operates with a relatively high water blowdown rate that would require high additive makeup rates. Therefore, increased venturi pressure drop and increased sulfite concentration (obtained by increasing the concentration of thiosulfate additive used to inhibit oxidation) were the upgrades tested at Elrama. Sulfite is an alkaline species with respect to SO₂ absorption, so increased sulfite levels were investigated as a lower cost alternative to using organic acid additives to increase liquid-phase alkalinity at this site.

Figure 3-6 shows an example of the effect of increased sulfite concentration at two levels of venturi pressure drop. NTU increases with increasing sulfite concentration until the sulfite concentration reaches about 2500 mg/L. Above this concentration the absorber performance is gas-film limited, and further increases in efficiency can only be obtained by increasing the effectiveness of gas-liquid contact. At Elrama, marginal improvements in gas-liquid contacting can be achieved by increasing the venturi pressure drop. Increasing the venturi pressure drop from 10 to 12 inches of water increased NTU by about 15 to 20%. The maximum SO₂ removal efficiency measured at Elrama was about 93%, limited primarily by the mass transfer area available in the venturi absorbers. This was the only site where the target performance of 95 to 98% SO₂ removal was not achieved in the single-module tests.
3.1.6 New York State Electric and Gas Corporation’s (NYSEG) Kintigh Station

NYSEG's Kintigh Station FGD system is an inhibited-oxidation limestone process with countercurrent-gas-flow spray tower absorbers. The absorber modules have five separate spray header elevations to provide gas/slurry contact area, each of which is supplied by an independent recycle pump. This configuration allows for convenient adjustment of SO$_2$ removal efficiency by adjusting the number of operating recycle pumps. Sodium formate additive was tested at Kintigh.

Figure 3-7 shows the effect of sodium formate additive at Kintigh for tests conducted at the normal pH set point of 5.6, a 9 ft/s flue gas velocity, and with either three, four, or five recycle pumps operating. With four pumps operating, the SO$_2$ removal efficiency increased from about 86% without additive to 99.4% with 3800 mg/L formate ion in the reaction tank. The shape of the performance curve suggests that the maximum gas-film-limited efficiency of the Kintigh absorbers is somewhat higher than this result. The tests using different numbers of operating slurry recycle pumps show that NTU was proportional to the slurry recirculation rate to the 0.7 power ($L^{0.7}$). Results from tests at varying flue gas velocities indicated that NTU was inversely proportional to the gas rate to the 0.25 power ($G^{-0.25}$).

3.2 Organic Acid Additive Consumption

Table 3-2 summarizes results of tests designed to measure the steady-state consumption rates of the organic acid additives (DBA and/or sodium formate) that were tested at five of the six sites. In these consumption tests, DBA or sodium formate was added continuously to the entire FGD system to maintain a fixed concentration, and the consumption rate was calculated by material balance.

There are several mechanisms by which these additives are consumed in FGD systems. With inhibited-oxidation processes, either additive can be lost by incorporation into the calcium sulfite byproduct solids. Neither additive precipitates with the gypsum byproduct in forced-oxidation processes, though. In either oxidation mode, additives are consumed by oxidative
## Table 3-2
Summary of Additive Consumption Test Results

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Upgrade Option</th>
<th>Test Duration (Hours)</th>
<th>Average Load (MW)</th>
<th>Concentration (mg/L DBA or formate ion)</th>
<th>Consumption (lb/hr DBA or sodium formate)</th>
<th>(lb DBA or sodium formate/ton SO2 removed)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tampa Electric Big Bend 4</td>
<td>DBA</td>
<td>90</td>
<td>400</td>
<td>320 (a) 640</td>
<td>60</td>
<td>6.5</td>
</tr>
<tr>
<td>Hoosier Energy Merom 2</td>
<td>DBA</td>
<td>312</td>
<td>360</td>
<td>1,200</td>
<td>120</td>
<td>9-10</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>139</td>
<td>325</td>
<td>2,500</td>
<td>170</td>
<td>23</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>43</td>
<td>325</td>
<td>2,900</td>
<td>340</td>
<td>36</td>
</tr>
<tr>
<td>Southwestern Electric Pirkey</td>
<td>DBA</td>
<td>70</td>
<td>520</td>
<td>1,100</td>
<td>110</td>
<td>11</td>
</tr>
<tr>
<td>PSI Energy Gibson 5</td>
<td>DBA</td>
<td>275</td>
<td>535</td>
<td>1,150</td>
<td>160</td>
<td>9</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>168</td>
<td>476</td>
<td>1,050</td>
<td>130</td>
<td>9</td>
</tr>
<tr>
<td>New York State Electric and Gas Corp. Kintigh</td>
<td>Sodium Formate</td>
<td>120</td>
<td>677</td>
<td>1,080</td>
<td>190</td>
<td>17</td>
</tr>
</tbody>
</table>

Notes: (a) For this dual-loop system, the upper-loop DBA concentration is less than the lower-loop concentration.
degradation and are lost with the liquor that remains adhered to the byproduct solids. Either additive would also be lost with any liquor blowdown from the FGD system. In addition, formate additive is lost by vaporization into the flue gas (as formic acid). DBA is not volatile at typical wet FGD process temperatures.

In Table 3-2, additive consumption rates are expressed as the average feed rate in lb/hr required to maintain the concentration that is shown, and as a "normalized" consumption rate per ton of SO₂ removed in the system. The consumption rate would be approximately proportional to the additive concentration within a specific FGD system. However, consumption rates can vary significantly from system to system depending on process chemistry.

The results show that DBA consumption rates were approximately the same (9 to 11 lb DBA/ton SO₂ removed at 1100 to 1200 mg/L) for the three different FGD systems in which similar concentrations were maintained (Merom, Pirkey, and Gibson) and was proportionately lower at Big Bend, where lower concentrations were used. Two of the three systems in which sodium formate was tested (Merom and Gibson) showed similar consumption rates for sodium formate (11 lb/ton SO₂ removed at 1050 mg/L at Gibson and an estimated 10 lb/ton SO₂ removed at 1100 mg/L at Merom), but the consumption rate at Kintigh was somewhat higher.

It is probably only coincidental that the organic acid additive consumption rates measured at four of the five sites where these additives were tested were very similar. The Kintigh results show that additive consumption rate can vary markedly among sites, depending on the oxidation mode and other process chemistry conditions.

3.3 Effects of Additives on Process Operation

While they are used primarily to enhance SO₂ removal efficiency, organic acid additives can potentially affect other process operating parameters in FGD systems, especially byproduct solids dewatering properties. During the additive tests at each site, potential side effects of additives were evaluated using several different measurement techniques. Table 3-3 summarizes observed effects of additives on process operation at each site.
Table 3-3
Effects of Additives on Process Operation

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Additive Tested</th>
<th>Settling Rate</th>
<th>Filtration Rate</th>
<th>Filter Cake % Solids</th>
<th>Sulfite Oxidation</th>
<th>Reagent Utilization</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tampa Electric Big Bend 4</td>
<td>DBA</td>
<td>Possible Decrease</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
</tr>
<tr>
<td>Hoosier Energy Merom 2</td>
<td>DBA</td>
<td>Increased</td>
<td>Not Tested</td>
<td>Not Tested</td>
<td>Possible Decrease</td>
<td>No Effect</td>
</tr>
<tr>
<td>Southwestern Electric Pirkey</td>
<td>Sodium Formate</td>
<td>No Effect</td>
<td>Not Tested</td>
<td>Not Tested</td>
<td>No Effect</td>
<td>No Effect</td>
</tr>
<tr>
<td></td>
<td>DBA</td>
<td>Increased</td>
<td>No Effect</td>
<td>No Effect</td>
<td>Decreased</td>
<td>Increased</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
<td>Decreased</td>
<td>Increased</td>
</tr>
<tr>
<td>PSI Energy Gibson 5</td>
<td>DBA</td>
<td>Increased</td>
<td>No Effect</td>
<td>Increased</td>
<td>No Effect</td>
<td>No Effect</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>Decreased</td>
<td>No Effect</td>
<td>Decreased</td>
<td>No Effect</td>
<td>No Effect</td>
</tr>
<tr>
<td>Duquesne Light Elrama</td>
<td>Thiosulfate</td>
<td>Increased</td>
<td>No Effect</td>
<td>No Effect</td>
<td>Decreased</td>
<td>No Effect</td>
</tr>
<tr>
<td>New York State Electric and Gas Corp. Kintigh</td>
<td>Sodium Formate</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
<td>No Effect</td>
</tr>
</tbody>
</table>
3.3.1 Settling Rate

Settling rate measurements showed that DBA significantly increased the settling rate of calcium sulfite solids in the inhibited oxidation processes at Merom, Pirkey, and Gibson. Scanning electron microscopy (SEM) confirmed that sulfite solids produced in the presence of DBA were thicker than the baseline solids at Merom and Pirkey, but no change in the solids appearance was seen at Gibson. DBA may have slightly decreased the settling rate of gypsum solids produced at Big Bend. SEM showed no detectable change in the size or shape of the gypsum solids produced in the presence of DBA at that site, though. Testing at ultra-high velocity conditions while using DBA additive showed no further effect on settling rates.

Sodium formate additive did not affect settling rates significantly at any site where it was tested, except for a possible decrease in settling rate at Gibson.

The settling tests also showed that settling rates vary widely during normal process operation at several of the systems. These changes appeared to be closely related to changes in boiler load, which change the extent of oxidation in the inhibited-oxidation processes and the solids residence time in the FGD reaction tank for either oxidation mode. The apparent changes in settling rate caused by the additives were within the range of these normal variations, so these systems had no problems dealing with the changes. For example, in the systems where DBA appeared to increase solids settling rates, thickener bed levels were lowered and these changes were successful in preventing problems such as excessive thickener rake torque.

3.3.2 Filtration Rate and Filter Cake Solids Content

Laboratory filtration tests yielded highly variable results with the slurries from each system. Filtration rates did not seem to be affected by the additives, and only in two systems was there a discernible effect on the final filter cake solids content. At Gibson, DBA was observed to increase the final filter cake solids content, while sodium formate decreased the final solids content. Increased thiosulfate concentration at Elrama decreased sulfite oxidation percentages as was expected, and also increased the final filter cake solids content in the laboratory filtration tests.
3.3.3 Sulfite Oxidation

Both DBA and sodium formate decreased sulfite oxidation at Pirkey. This is a favorable effect for inhibited oxidation systems. DBA may also have decreased oxidation at Merom, also a favorable effect there. At other systems, the organic acid additives had no observed effect on oxidation. Normal variations in oxidation with changes in boiler load were also observed, though.

At Elrama, increased thiosulfate concentration showed the expected effect of lowering sulfite oxidation percentages.

3.3.4 Reagent Utilization

The additives had no effect on reagent utilization with the exception of a moderate increase in limestone utilization observed at Pirkey.
4.0 FGDPRISM SIMULATIONS

The FGDPRISM computer model developed by the Electric Power Research Institute was calibrated for each of the six systems using the physical dimensions of each absorber and the data from the baseline and parametric tests. Performance predictions using the calibrated models were used along with actual test data to evaluate the economics of potential upgrade options. The computer model was used to examine some options that could not be conveniently tested and for some sites to normalize test results to a common set of conditions.

4.1 FGDPRISM Calibration

The model was calibrated to each specific FGD system by adjusting various parameters so that predicted \( \text{SO}_2 \) removal efficiency matched the measured efficiency at similar levels of slurry \( \text{pH} \), reagent utilization, sulfite oxidation percentage, and process liquor composition. Test data that represented a wide range of \( \text{SO}_2 \) removal and reagent utilization values were used wherever possible to provide the best calibration results.

Figure 4-1 shows an example of predicted versus actual \( \text{SO}_2 \) removal efficiency for the calibration cases from the first series of tests (normal velocity) at TECo's Big Bend station. Figure 4-2 shows how well the calibrated model predicted \( \text{SO}_2 \) removal for other tests at Big Bend that were not used as calibration cases. The maximum deviation between measured and predicted efficiency for cases not used in the calibration was about three percentage points.

4.2 FGDPRISM Performance Predictions

The calibrated FGDPRISM models for the six test sites were used to evaluate a variety of performance upgrade options that were not actually tested. Extrapolation of test results beyond the range of test conditions is a useful application of this computer model. Table 4-1 shows how the calibrated FGDPRISM models were used at the different sites to extend the usefulness of test results. These applications are discussed briefly below.
Figure 4-1. Comparison of Observed and Predicted Values for Overall SO\(_2\) Removal for the Calibration Cases at Tampa Electric Big Bend Unit 4

Figure 4-2. Comparison of Observed and Predicted Values for Overall SO\(_2\) Removal for Cases Not Used in the Calibration for Tampa Electric Big Bend Unit 4
<table>
<thead>
<tr>
<th>Test Site</th>
<th>Upgrade Options Tested</th>
<th>Upgrade Options Evaluated Using FGDPRISM</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tampa Electric Big Bend 4</td>
<td>DBA</td>
<td>Increased or decreased packing depth with various DBA concentrations.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Increased or decreased L/G with various DBA concentrations.</td>
</tr>
<tr>
<td>Hoosier Energy Merom 2</td>
<td>DBA Sodium Formate</td>
<td>Increased or decreased packing depth with various DBA and formate concentrations</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Closed-loop operation with additives.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Operation with higher sulfur fuel.</td>
</tr>
<tr>
<td>Southwestern Electric Pirkey</td>
<td>DBA Sodium Formate</td>
<td>Use of DBA instead of excess reagent to maintain compliance at very high fuel sulfur levels.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Operation with higher sulfur fuel.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Increased dissolved magnesium levels.</td>
</tr>
<tr>
<td>PSI Energy Gibson 5</td>
<td>DBA Sodium Formate</td>
<td>Comparison of baseline and upgrade results at consistent levels of inlet SO₂ and limestone grind.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Operation with higher sulfur fuel.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Increased dissolved magnesium levels.</td>
</tr>
<tr>
<td>Duquesne Light Elrama</td>
<td>Venturi ΔP Thiosulfate</td>
<td>Adding packing to the venturi absorber.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Operation with DBA additive.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Operation at higher magnesium concentrations.</td>
</tr>
<tr>
<td>New York State Electric and Gas Corp Kintigh</td>
<td>Sodium Formate</td>
<td>Operation with a finer limestone grind.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Use of sodium formate to maintain current SO₂ removal with fewer recycle pumps.</td>
</tr>
</tbody>
</table>
Changes in packing depth at TECo’s Big Bend station and at Hoosier Energy’s Merom station were evaluated using FGDPRISM. In these packed absorbers, SO₂ removal efficiency is a strong function of packing depth, but several days of down time and significant materials and labor costs would have been incurred to actually test different packing depths. FGDPRISM predictions used in the economic evaluations showed that increased packing depth would be a cost-effective upgrade for both of these FGD systems, provided that increased maintenance costs did not result because of increased plugging which could potentially occur with greater packing depth.

SO₂ removal is also strongly affected by the absorber L/G. At TECo’s Big Bend FGD system, multiple recycle pumps are connected to common spray headers. Because of the flow characteristics of centrifugal pumps and pressure spray nozzles, changing the number of operating pumps with this configuration results in only small changes in L/G and also causes the spray nozzles to operate outside of their intended pressure range. However, changes in L/G could be implemented, if desired, by changing the spray nozzles along with changing the number of operating pumps. For this system, the cost effectiveness of operation at higher or lower L/G was evaluated using FGDPRISM. The FGDPRISM results were subsequently used in economic evaluations, which showed no advantage for the high or low L/G cases relative to normal operation.

The effect of changes in fuel sulfur content can also be modeled with FGDPRISM. For example, at SWEPCo’s Pirkey Station, the lignite heating value and sulfur content are highly variable. During the performance tests, the inlet SO₂ ranged from 3.0 to 5.1 lb SO₂/million Btu, but levels as high as 8 lb SO₂/million Btu are occasionally encountered. The calibrated model for the Pirkey system was used to evaluate the cost effectiveness of using DBA additive instead of very high limestone reagent ratios (high excess reagent levels in the recirculating slurry solids) to maintain emissions compliance at high inlet SO₂ levels. The model results and subsequent economic evaluations showed the DBA additive would be an effective means of maintaining compliance under such high inlet SO₂ conditions.

Another important process performance variable is limestone grind fineness. At NYSEG’s Kintigh Station, the current limestone grind is relatively coarse. A finer grind could be
produced by operating the limestone ball mills at a lower throughput for a longer period of time, but modifications to the classifiers (size separation devices) in the grinding circuit would be required. FGDPRISM was used to evaluate the potential performance improvement that could be obtained with a finer limestone grind. The results suggested that this would be a promising upgrade option.

FGDPRISM also predicts consumption rates for sodium formate and DBA additives. Table 4-2 compares measured and predicted additive consumption rates for four of the five sites where these additives were tested. FGDPRISM could not predict DBA consumption for the TECO dual-loop system because of widely different chemistry conditions in the two loops. These results show that the predicted consumption rates are all within about 25% of the measured rates except for the second formate test at Merom, where the predicted rate was only 58% of the measured value.

Table 4-2
Predicted Versus Measured Additive Consumption

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Additive Tested</th>
<th>Average Concentration (mg/L)</th>
<th>Measured Consumption (lb additive/ton SO₂ Removed)</th>
<th>FGDPRISM Predicted Consumption (lb additive/ton SO₂ Removed)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hoosier Energy Merom 2</td>
<td>DBA</td>
<td>1,200</td>
<td>10</td>
<td>11</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>2,500</td>
<td>23</td>
<td>18</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>2,900</td>
<td>36</td>
<td>21</td>
</tr>
<tr>
<td>Southwestern Electric Pirkey</td>
<td>DBA</td>
<td>1,170</td>
<td>11</td>
<td>12</td>
</tr>
<tr>
<td>PSI Energy Gibson 5</td>
<td>DBA</td>
<td>1,150</td>
<td>9</td>
<td>7</td>
</tr>
<tr>
<td></td>
<td>Sodium Formate</td>
<td>1,050</td>
<td>11</td>
<td>13</td>
</tr>
<tr>
<td>New York State Electric and Gas Corp. Kintigh</td>
<td>Sodium Formate</td>
<td>1,080</td>
<td>17</td>
<td>13</td>
</tr>
</tbody>
</table>
5.0 ECONOMIC EVALUATIONS

Test results and FGDPRISM performance predictions were used along with pertinent cost data to evaluate the cost effectiveness of SO$_2$ removal upgrade options at each of the six sites. Cost data were provided by the host utilities and by additive vendors. All of the upgrade options were evaluated over an appropriate range of operating conditions (e.g., slurry pH, L/G, additive concentration) to determine the optimum conditions. Table 5-1 summarizes the results of the economic evaluations. In this table, only the most cost-effective upgrades are shown. Results for the individual sites are discussed briefly below.

5.1 Tampa Electric Company’s Big Bend Unit 4

The baseline module efficiency at Big Bend was about 93%. The FGD system efficiency could be increased to about 99% by adding 500 mg/L DBA to the system at the normal flue gas velocity of 7.5 ft/s. Under these conditions, treating only flue gas from Unit 4, an additional 4,400 tons/yr of SO$_2$ could be removed at an incremental cost of about $65/ton. This cost includes the annualized cost of a new additive storage and injection system plus the cost of additional limestone reagent and the DBA, at $0.21 per lb dry weight. A small credit was taken for additional gypsum byproduct revenue. The net annual value of this upgrade is about $400,000, assuming an SO$_2$ allowance value of $150/ton.

FGDPRISM predictions showed that increasing the packing depth to three feet from the normal two feet in the upper absorber loop would also be a cost-effective upgrade at Big Bend, but this option was judged by TECO to be unattractive from an operational standpoint because of the potential for increased plugging.

As described in Section 3, the testing at Big Bend also showed that DBA could be used to maintain high SO$_2$ removal efficiencies while operating the test absorber at velocities about 33% higher than the original design. Based on these and subsequent encouraging test results, TECO converted the existing Unit 4 FGD system to scrub flue gas from the adjacent Unit 3. Since 1995, they add 1000 mg/L of DBA to the upper loop and operate all four absorbers (no spare) at flue gas
### Table 5-1
Results of Economic Evaluations of SO₂ Removal Upgrade Options

<table>
<thead>
<tr>
<th>Test Site</th>
<th>Baseline SO₂ Removal Efficiency</th>
<th>Upgrade Option</th>
<th>Optimum SO₂ Removal Efficiency</th>
<th>Additional SO₂ Removed (ton/year)</th>
<th>Total Cost of Additional SO₂ Removal</th>
<th>Value of Additional SO₂ Removal (a)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tampa Electric Co. Big Bend 4</td>
<td>93</td>
<td>500 mg/L DBA</td>
<td>99%</td>
<td>4,400</td>
<td>$287,000</td>
<td>65</td>
</tr>
<tr>
<td>Hoosier Energy Merom 2</td>
<td>83 (b)</td>
<td>Close bypass 2,800 mg/L DBA</td>
<td>97%</td>
<td>15,100</td>
<td>$833,000</td>
<td>55</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Close bypass 2,500 mg/L formate</td>
<td>95%</td>
<td>12,900</td>
<td>$799,000</td>
<td>62</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Close bypass Increase reagent ratio</td>
<td>94%</td>
<td>11,500</td>
<td>$527,000</td>
<td>46</td>
</tr>
<tr>
<td>Southwestern Electric Co. Pirkey</td>
<td>80 (b)</td>
<td>Close bypass</td>
<td>98%</td>
<td>20,200</td>
<td>$841,000</td>
<td>42</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Close bypass 1,000 mg/L DBA</td>
<td>99%</td>
<td>21,200</td>
<td>$822,000</td>
<td>39</td>
</tr>
<tr>
<td>PSI Energy Gibson 5</td>
<td>80 (b)</td>
<td>Close bypass 1,200 mg/L DBA</td>
<td>99%</td>
<td>12,900</td>
<td>$640,000</td>
<td>48</td>
</tr>
<tr>
<td></td>
<td>Close bypass Operate spare module</td>
<td>93%</td>
<td>13,200</td>
<td>$640,000</td>
<td>48</td>
<td>$1,300,000</td>
</tr>
<tr>
<td></td>
<td>Close bypass Operate spare module</td>
<td>95%</td>
<td>15,400</td>
<td>$910,000</td>
<td>59</td>
<td>$1,400,000</td>
</tr>
<tr>
<td></td>
<td>Increase reagent ratio</td>
<td>95%</td>
<td>15,400</td>
<td>$910,000</td>
<td>59</td>
<td>$1,400,000</td>
</tr>
<tr>
<td></td>
<td>Close bypass Operate spare module</td>
<td>95%</td>
<td>14,800</td>
<td>$910,000</td>
<td>62</td>
<td>$1,300,000</td>
</tr>
<tr>
<td></td>
<td>Increase reagent ratio</td>
<td>95%</td>
<td>15,400</td>
<td>$910,000</td>
<td>59</td>
<td>$1,400,000</td>
</tr>
<tr>
<td></td>
<td>Operate spare module 250 mg/L formate</td>
<td>95%</td>
<td>15,400</td>
<td>$910,000</td>
<td>59</td>
<td>$1,400,000</td>
</tr>
<tr>
<td></td>
<td>Operate spare module 500 mg/L DBA</td>
<td>95%</td>
<td>15,400</td>
<td>$910,000</td>
<td>59</td>
<td>$1,400,000</td>
</tr>
<tr>
<td>Duquesne Light Co. Elrama</td>
<td>89</td>
<td>Increase venturi DP Decrease reagent ratio</td>
<td>92%</td>
<td>1,480</td>
<td>($138,000)</td>
<td>(94)</td>
</tr>
<tr>
<td></td>
<td>Increase thiosulfate</td>
<td>92%</td>
<td>1,480</td>
<td>($138,000)</td>
<td>(94)</td>
<td>$360,000</td>
</tr>
<tr>
<td>New York State Electric and Gas Corp. Kintigh</td>
<td>85 (b)</td>
<td>Close bypass 1000 mg/L formate</td>
<td>98%</td>
<td>10,600</td>
<td>$810,000</td>
<td>76</td>
</tr>
</tbody>
</table>

Notes:  
- a – SO₂ allowance value assumed to be $150/ton.  
- b – Baseline performance includes partial flue gas bypass.
velocities about 50% greater than the design value, to treat flue gas from both the 485-MW Unit 4 and 445 MW Unit 3 in the Unit 4 FGD system at Big Bend.

This modification to double the amount of flue gas treated in the system was more capital intensive than was the intent of this project, so we did not estimate the cost effectiveness of this conversion. Furthermore, TECo did not wish to publish the specifics of the cost effectiveness of this implementation, either relative to the cost of a new FGD system for Unit 3 or on a dollars per additional ton of SO₂ removed basis. However, TECo’s modifications at Big Bend demonstrate another potentially cost-effective approach to high-efficiency operation of an existing FGD system: using additives to allow the existing system to treat more flue gas.

Similarly, after the 1997 ultra-high velocity testing at Big Bend, it was beyond the scope of this project to estimate the costs to convert the Unit 4 FGD system to treat flue gas from a third unit. However, the fact that TECo was willing to make several significant modifications to one absorber to host this 1997 testing demonstrates the potential they see for this approach to achieving compliance with Phase II of the 1990 Clean Air Act amendments.

5.2 **Hoosier Energy’s Merom Unit 2**

The Merom FGD system normally operates with partial flue gas bypass at an overall efficiency of about 83%. DBA was the most cost-effective additive at Merom. The FGD system efficiency could be increased to about 97% by closing the bypass and adding 2,800 mg/L of DBA. With this option, an additional 15,000 tons/yr of SO₂ could be removed at an average incremental cost of $55/ton. Sodium formate was slightly less effective, yielding an optimum SO₂ removal efficiency of 95%, with about 13,000 tons/yr additional SO₂ removal at an average cost of $62/ton.

Increasing the reagent ratio (increased excess reagent in the recirculating slurry) without additive also provides a substantial increase in SO₂ removal at Merom, yielding an SO₂ removal efficiency of about 94% and removing 11,500 tons/yr additional SO₂ removal at an average cost of only $46/ton. The net annual value of these upgrades ranged from about $1.2 million to $1.4 million assuming an SO₂ allowance value of $150/ton.
FGDPRISM predicted that increased packing depth would also be cost-effective for the Merom system, reducing the required DBA concentration to obtain 97% SO₂ removal and increasing the net annual value of additional SO₂ removed by about 10%.

5.3 **Southwestern Electric Power Company’s Pirkey Station**

The baseline FGD system efficiency at Pirkey was about 80%, because nearly 20% of the boiler flue gas was bypassed around the FGD absorbers. At Pirkey, the absorber efficiency was high without additive, and a significant amount of additional SO₂ could be removed simply by closing the bypass damper. A system efficiency of 98% was estimated for this option, removing more than 20,000 ton/yr of additional SO₂ at an average cost of $42/ton. The net annual value of this option was about $2.2 million, based on an SO₂ allowance value of $150/ton. The system efficiency could be increased to 99% with 1000 mg/L of DBA additive, removing an additional 1,000 ton/yr of SO₂ and increasing the net annual value of allowances by nearly $200,000 to about $2.4 million.

The Pirkey results also showed that DBA addition at Pirkey could be cost effective even without decreasing the amount of flue gas bypass and increasing the system SO₂ removal. DBA addition was observed to have the effect of lowering oxidation percentages in this inhibited-oxidation system to consistently below 15%, thus reducing the potential for gypsum-scale formation in the absorbers. Also, DBA addition could allow lower recirculating slurry pH set points and a correspondingly higher limestone reagent utilization in the system. Subsequent to this testing, SWEPCo installed a DBA additive system at Pirkey, and is currently realizing these observed operating and maintenance cost savings while maintaining their currently required average of approximately 80% overall SO₂ removal.

5.4 **PSI Energy’s Gibson Unit 5**

At Gibson, the FGD system is operated with partial flue gas bypass to obtain about 80% SO₂ removal efficiency. The system efficiency could be increased to about 93% by closing the bypass and operating the spare module. This option removes more than 13,000 ton/yr of additional SO₂ at an average cost of $48/ton, for a net annual value of $1.3 million. The system
efficiency can be increased to about 95% using one of three options: increasing the reagent ratio (higher excess reagent levels in the recirculating slurry solids), using DBA additive, or using sodium formate additive. The DBA option and higher reagent ratio increase the net annual value by about $100,000. The use of additives at Gibson would probably not be justified unless the value of allowances increases beyond the assumed value of $150/ton. Although even this value is well above the average price of SO$_2$ allowances in recent EPA auctions, allowances are expected to increase in value during Phase II of the Clean Air Act Amendments, beginning in the year 2000.

5.5  **Duquesne Light Company's Elrama Station**

This system is a magnesium-enhanced lime process that is already operating close to the maximum SO$_2$ removal efficiency that can be obtained without major equipment modifications. The baseline efficiency of 86 to 89% could be increased to 92% by increasing the amount of thiosulfate used to inhibit oxidation, thus increasing liquid-phase sulfite concentrations, and by operating the venturi absorber modules at higher pressure drop to increase mass transfer surface area. Under these conditions, the slurry pH set point can be decreased, resulting in a net operating savings of $94/ton of additional SO$_2$ removed. Only about 1,500 ton/yr of additional SO$_2$ removal can be obtained at Elrama, but the significant savings in lime consumption by operating in this mode brings the net annual value to nearly $400,000.

5.6  **New York State Electric and Gas Corporation’s Kintigh Station**

Sodium formate additive was the primary upgrade option considered for the Kintigh system. Sodium formate additive at 1000 mg/L was estimated to yield a cost-effective increase in SO$_2$ removal efficiency from the baseline value of 85% to 98%. More than 10,000 ton/yr of additional SO$_2$ could be removed at an average cost of $76/ton. The net annual value of this upgrade is about $800,000.

Performance predictions using FGDPRISM also showed that use of a finer limestone grind at Kintigh could yield a cost savings of $200,000/yr while operating at the current emissions limit.
6.0 REFERENCES


