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High SO₂ Removal Efficiency Testing

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Topical Report

Results of DBA Additive Tests at Tampa Electric Company's Big Bend 4 FGD System

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SUMMARY

Tests were conducted on the Tampa Electric Company's (TECo) Big Bend Station, Unit 4 flue gas desulfurization (FGD) system to evaluate an option for achieving high sulfur dioxide (SO₂) removal efficiencies. The option tested was the addition of dibasic acid (DBA) additive. In addition, the effectiveness of other potential options was simulated with the Electric Power Research Institute's (EPRI) FGD Process Integration and Simulation Model (FGDPRISM) after it was calibrated to the system. An economic analysis was conducted to determine the cost effectiveness of each option considered. The following summarizes these results.

SO₂ Removal Performance. Baseline tests on one absorber module of the Unit 4 double-loop wet limestone FGD system showed that, at normal operating conditions, the SO₂ removal efficiency of the module was about 93%. The normal operating conditions include forced sulfite oxidation, pH set points of 5.7 to 5.8 in the upper loop and 4.2 in the lower loop, a superficial flue gas velocity of 7.5 ft/s through the absorber, two feet of packing in the upper-loop section, and a total liquid-to-gas ratio (L/G) for the two loops of 84 gal/macf.

The module SO₂ removal efficiency could be increased to 97+% by increasing the pH set points to 6.2 in the upper loop and 5.2 in the lower loop, but these set points do not represent feasible long-term operating conditions. High-pH operation in the upper loop will tend to increase plugging in the packing and mist eliminator. High pH set points in the lower loop will not allow the production of a wallboard-grade gypsum byproduct, because the excess limestone results in an unacceptable gypsum purity level.

Tests with DBA additive showed that the module SO₂ removal efficiency could be increased to as high as 99.7% at an upper-loop DBA concentration of about 1000 ppm, with the normal pH set point in the upper loop and a lower, 3.8 set point in the lower loop. An upper-loop DBA concentration of only 250 to 300 ppm was required to achieve 98% SO₂ removal. DBA addition appeared to have no significant effects on process operation or gypsum solids properties,

with the possible exception of a slight decrease in gypsum byproduct filter cake solids content during this testing. This decrease might have been due to other effects, though, such as a buildup of fines within the FGD system that was unrelated to DBA addition.

DBA Additive Consumption. A long-term DBA consumption test was conducted where DBA was added to the entire FGD system. At an average unit load of 400 MW, a DBA feed rate of about 59 lb/hr was required to maintain a DBA concentration of 320 ppm in the upper-loop reaction tanks. This concentration was selected to achieve at least 98% SO₂ removal across the FGD system. However, the test-average SO₂ removal was actually only slightly greater than 95%. It appears that this lower average resulted because one of the three operating modules contained a less efficient packing configuration, and was likely only achieving 90% SO₂ removal while the other two modules were achieving 98% removal.

About 83% of the total DBA consumption was due to degradation (i.e., losses other than with liquor leaving the FGD system). On an SO₂ removal basis, the measured DBA degradation rate was 5.4 ± 1 lb per ton of SO₂ removed. The total DBA consumption rate, including solution losses, was 6.5 lb per ton of SO₂ removed.

SO₂ Removal Upgrade Economics. The economics of DBA addition were evaluated based on additive feed system capital costs estimated by Stone & Webster Engineering Corporation, operating costs provided by TECo, and delivered DBA costs provided by Monsanto (\$0.21/lb) and DuPont (\$0.33/lb). These estimates show the economics of DBA addition to achieve high SO₂ removal efficiencies to be very favorable.

At the lower DBA cost, the net cost for additional tons of SO₂ removed by DBA addition was estimated to remain below \$100/ton until over 99% removal efficiency is obtained. The maximum net annual value of the additional SO₂ removed with DBA addition is about \$820,000, if allowances are valued at \$250/ton. If the allowances are instead valued at a more conservative \$150/ton, the maximum net annual value is still about \$380,000.

Results of FGDPRISM simulations were used to evaluate upgrade options other than those tested. Operation with one or three upper-loop slurry recycle pumps (versus the two normally operated) and with one or three feet of packing in the upper loop (versus the normal packing depth of two feet) were evaluated. Increasing the number of operating slurry recycle pumps was not a favorable option, nor was reducing the packing depth to one foot.

Operation with one slurry recycle pump appeared to be cost effective, but would require as much as four times the DBA consumption rate as operating with two pumps. The one-pump option was therefore sensitive to DBA cost, and could also result in operating problems that were not considered in the economic analysis. Operation with increased packing depth (with or without DBA addition) yielded economic results comparable to those of the DBA-only option. However, this option also could lead to operating problems, such as increased plugging in the packed section of the modules, and could therefore incur maintenance costs that were not considered in this evaluation. Consequently, DBA addition in the normal operating configuration (two feet of packing and with two upper-loop slurry recirculation pumps in service) appears to be the best approach for achieving high-efficiency SO_2 removal operation with the Big Bend Unit 4 FGD system.

A goal of this project of achieving 95 to 98% overall SO_2 removal with an existing FGD system designed to achieve 90% removal appears to be readily exceeded at this site. Adding DBA to a concentration of 500 ppm in the upper-loop slurry should allow the Big Bend Unit 4 FGD system to achieve over 99% SO_2 removal. This would result in the removal of an additional 4400 tons/yr of SO_2 , at an estimated cost of only \$65 per additional ton removed. This approach is very attractive compared to the anticipated costs of purchasing SO_2 allowances in the marketplace, or costs for retrofitting SO_2 removal capacity on unscrubbed units.

1.0 INTRODUCTION

This report describes the results of tests conducted on Tampa Electric Company's (TECo) Big Bend Station Unit 4 flue gas desulfurization (FGD) system. The objective of these tests was to obtain full-scale data to evaluate the cost effectiveness of upgrading the SO₂ removal performance of an existing FGD system, as part of a strategy for meeting the requirements of the 1990 Clean Air Act (CAA) amendments. The upgrade tested at this site was the use of dibasic acid, an organic acid buffer, as a performance additive.

1.1 Background

Provisions of the CAA amendments call for a ten-million ton per year reduction in U.S. SO₂ emissions (from a 1980 baseline) in two phases. Phase I calls for a five-million ton per year reduction 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, installing new 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 (FGC) Program at the U.S. Department of Energy Pittsburgh Energy Technology Center (DOE PETC) 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 particulates, and air toxics generated from coal combustion. A portion of the FGC Program, including this project, involves enhancing the SO₂ removal efficiencies 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 I and/or Phase II of the Clean Air Act amendments.

In this project, Radian Corporation is conducting tests at six full-scale FGD systems to evaluate options for achieving high SO₂ removal efficiencies (95 to 98% removal). Each system is being characterized under baseline operation, and then with additives or with other modifications to enhance SO₂ removal performance.

The systems being evaluated are at the Tampa Electric Big Bend Station, the Hoosier Energy Merom Station, the Southwestern Electric Power Company Pirkey Station, the PSI Energy Gibson Station, the Duquesne Light Elrama Station, and the New York State Electric and Gas Company Kintigh Station. A wide variety of FGD system vendors and designs are represented in the program. Most of these systems were designed to achieve 85 to 90% SO₂ removal.

This Topical Report only covers results from the first site, at Tampa Electric's Big Bend Station. Separate Topical Reports will be issued for each subsequent site.

1.2 Project Description

Three types of performance tests were completed at Big Bend. First, "baseline" tests were conducted to obtain performance data without the DBA additive. Then, "parametric" tests were done to obtain performance data using DBA additive at various concentrations. The baseline and parametric tests were conducted using only one of the three operating scrubber modules. Following the parametric tests, a steady-state DBA consumption test was conducted, during which DBA was added to the entire FGD system.

Under a separate project funded by the Electric Power Research Institute, their FGD Process Integration and Simulation Model (FGDPRISM) was calibrated to the Big Bend Unit 4 FGD system. The calibrated model was then used to predict system SO₂ removal performance over a range of process conditions and DBA concentrations.

Economic calculations were conducted to determine the most cost-effective approach for achieving the project target of 95 to 98% SO₂ removal with the Big Bend Unit 4 FGD system. Actual and predicted performance results, along with measured steady-state DBA consumption data, additive system capital cost estimates by Stone and Webster Engineering Corporation, and other pertinent cost information provided by Tampa Electric Company, provided the basis for the economic evaluation. The net cost for additional tons of SO₂ removal was estimated for different operating conditions and DBA concentrations. 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 operating conditions for this system from the standpoint of compliance with the CAA amendments.

1.3 Report Organization

The performance tests are described and results are presented and discussed in Section 2. The FGDPRISM calibration procedure and performance predictions are discussed in Section 3, and the economic evaluation is detailed in the final section of this report, Section 4.

2.0

FULL-SCALE TESTING DESCRIPTION AND RESULTS

This section describes full-scale FGD system testing conducted at the Tampa Electric Company (TECo) Big Bend Station Unit 4, and provides an overview and discussion of the results. Detailed results are tabulated in Appendices A, B, C, and D. The full-scale tests were conducted to evaluate methods of achieving high SO₂ removal efficiencies with the Big Bend Unit 4 FGD system, and followed a methodology that has been used for other sites included in this DOE-PETC program.

The testing started with baseline tests on a single module of the FGD system. This established the "as found" performance of the system, and allowed collection of enough information to calibrate the EPRI FGDPRISM model to the FGD system. The calibrated FGDPRISM model was then used to evaluate potential SO₂ removal upgrade options.

Next, a series of short-term parametric tests was conducted, also on a single module, to verify model predictions and to demonstrate the SO₂ removal capabilities with the upgrade option in place. For this site, the only upgrade option tested at full scale was the addition of DBA, an organic acid buffer.

Finally, an additive consumption test was performed on the entire FGD system for a duration of approximately 90 hours. This test was conducted to measure the DBA addition rate required to maintain the target concentration for achieving high SO₂ removal efficiency. This allowed the costs of this upgrade approach to be more accurately determined, and demonstrated high-efficiency SO₂ removal operation with the entire FGD system rather than with only a single module.

This section presents and discusses results from each of these three phases of testing at the Big Bend site. In Section 2.1, there is a description of the Unit 4 FGD system. The testing approach and measurement methods are described in Sections 2.2 and 2.3, respectively. The remainder of the section presents results from the test program. Results for the baseline test

period are presented and discussed in Section 2.4, results for the DBA additive parametric test period are presented and discussed in Section 2.5, and results from the system-wide DBA consumption test are discussed in Section 2.6. Finally, Section 2.7 provides a discussion of the observed effects of DBA addition on FGD byproduct solids dewatering properties, for both the parametric and DBA consumption test period.

2.1 FGD System Description

Unit 4 at the Tampa Electric Big Bend Station is rated at 455 MW, and fires a 3% sulfur coal. The FGD system was designed to achieve 90% SO₂ removal at 97% limestone reagent utilization, and operates in a forced-sulfite-oxidation mode. A wallboard-grade gypsum byproduct is produced.

Flue gas leaving the Unit 4 boiler is treated in an ESP for particulate removal before entering the FGD system. Three identical 36-ft diameter modules treat the flue gas. A fourth module remains out of service as a spare. Each of the four absorbers has its own upstream booster fan.

Figure 2-1 illustrates the arrangement of a single scrubber module. The scrubber modules are Research-Cottrell Double-Loop™ units modified by Tampa Electric to feed limestone and oxidation air independently to each of the two loops. The original design fed limestone only to the upper loop, and oxidation air only to the lower loop.

The lower or quench loop operates at a relatively low pH and saturates the flue gas while removing HCl. The low pH set point promotes complete sulfite oxidation and high limestone utilization, which are required to produce wallboard-grade gypsum. A portion of the flue gas SO₂ is also removed in the lower loop. The slurry flow rate to the lower-loop spray headers corresponds to a liquid-to-gas ratio (L/G) of about 21

Figure 2-1. TECO Big Bend Unit 4 Scrubber Module with Sampling Locations

gal/kacf, based on measurements made during this program. The reaction tank for the lower-loop recirculating slurry is located in the base of the absorber.

Flue gas leaving the lower loop flows around a bowl-type slurry collector into the upper loop. The bowl collects slurry from the upper loop to return to the reaction tank, while allowing the flue gas to flow from one loop to the other. The upper loop operates at a higher recirculating slurry pH set point, and has a packed section to promote high SO₂ removal efficiency. There is one slurry flooding header above the packing and two conventional spray headers below the packing. The total upper-loop L/G was measured at about 64 gal/kacf, at the normal full-load superficial flue gas velocity of 7.5 ft/sec through the absorber. Each absorber upper loop has a separate, external reaction tank.

As the flue gas exits the upper loop, it is treated in a vertical-gas-flow, two-stage chevron mist eliminator, then exits the absorber. The gas from the three operating absorbers is combined, and then goes to the Unit 4 stack. Heated air is mixed with the saturated flue gas to provide reheat before the gas enters the stack.

Slurry blowdown to dewatering is only from the lower loop of each operating absorber. Primary dewatering is by hydroclones, and secondary dewatering is by a rotary drum vacuum filter. The filter cake is washed with fresh water to remove chlorides to ensure that the gypsum byproduct meets wallboard specifications.

The chloride-containing water recovered in the primary and secondary dewatering devices is returned to the lower loop. Only fresh water is made up to the upper loop, primarily through mist eliminator washing and with the limestone reagent slurry makeup. By avoiding the use of chloride-rich return water in the upper loop, lower alloy materials of construction can be used there.

There is a continuous blowdown of liquor from the FGD system to control the concentration of chloride in the recirculating liquor. This blowdown also serves to remove fine

gypsum particles from the system, which otherwise would build up with time because they are not effectively removed in the hydroclone primary dewatering devices.

2.2 Test Approach

As described in the introduction to this section, the baseline and parametric tests were conducted on a single module (Module B). Flue gas and slurry sampling points as well as DBA addition points around the test module are indicated in Figure 2-1. The DBA consumption test was conducted on all three operating modules.

2.2.1 Baseline Tests

For the baseline tests, independent variables included upper- and lower-loop slurry pH set points, flue gas velocity, and the number of upper-loop slurry recirculation pumps operating. The upper- and lower-loop pH set points, and the number of upper-loop slurry recirculation pumps operating could be varied independent of the operation of the other two modules in service. Holding the flue gas volume treated by Module B constant at a specified operating condition did impact the other two operating modules. This was done by holding the Module B booster fan inlet vane position constant, then varying the flue gas flow to the other two modules to maintain a constant inlet duct pressure. This allowed the Module B test conditions to be maintained independent of boiler load.

Flue gas velocity was of particular interest for this site because an unscrubbed unit is located adjacent to Unit 4. One part of the Clean Air Act amendment compliance strategy for Tampa Electric could be to add ductwork to allow all or part of the flue gas from the adjacent unit to be scrubbed in the Unit 4 FGD system. To evaluate this possibility, some of the tests were conducted at the maximum flue gas velocity achievable through the test module, which proved to be about 45% higher than the normal full-load velocity.

The first baseline test was conducted at TECo's normal operating conditions, over a period of two days. This test duration was adequate to approach steady-state conditions with respect to solids properties in the test module. For this two-day test, performance indicators included SO₂ removal efficiency, limestone utilization, sulfite oxidation percentage, relative saturations of limited-solubility species, and solids dewatering properties.

The remaining baseline tests were of half-day duration. For these shorter tests, SO₂ removal efficiency and limestone utilization were the primary performance indicators. Solids properties were not expected to reach steady-state conditions during these tests. The conditions for the half-day baseline tests were chosen to represent extremes of the operating range of interest. Measurements of SO₂ removal efficiency and limestone utilization over a wide pH range are used for calibrating FGDPRISM. The actual test conditions and SO₂ removal results are discussed later in this section.

2.2.2 Parametric Tests

For the parametric tests, independent variables were the same as for the baseline tests, with slurry DBA concentration as an additional variable. The parametric tests were arranged in groups of two to five tests at increasing levels of DBA concentration.

The first parametric test at each DBA level was a two-day test, so the effect of increasing DBA concentration on solids properties could be evaluated. All of the other tests were one-day or half-day tests for which the major performance indicators were SO₂ removal efficiency and limestone utilization.

Special provisions had to be made to control the DBA concentrations in the upper and lower loops of Module B. As mentioned above, fresh makeup water is added to the upper loop of the module through the mist eliminator wash and with the limestone reagent slurry, and water returned from the dewatering system is added only to the lower loop. Slurry from the upper-loop reaction tank overflows into the lower-loop reaction tank. With this configuration,

DBA would normally be added only to the upper-loop slurry, as most of the SO_2 removal occurs in that loop. DBA would reach the lower-loop slurry only with the overflow from the upper loop and with process water returned from the dewatering area.

Because the parametric tests were conducted on a single module, DBA concentrations did not reach steady-state levels in other portions of the FGD system. To simulate steady-state conditions in the test module, the expected steady-state distribution of DBA between the upper and lower loops was estimated by material balance calculations. To simulate this distribution during the short-term parametric tests, DBA was added to both loops. As the tests proceeded, the DBA concentration in the process water returning to the lower loop gradually increased. This concentration was measured daily, and the DBA feed rate to the lower loop was adjusted accordingly.

2.2.3 Additive Consumption Test

Following completion of the parametric tests, the upper-loop DBA concentration necessary to obtain 98% SO_2 removal was determined from the parametric test data. The DBA feed rate required to maintain that concentration was estimated using a system material balance model and parametric test data. During the consumption test, Modules B, C, and D were operated at the normal upper-loop pH, but with a reduced lower-loop pH set point to limit the amount of SO_2 removed there with DBA present. This was to avoid having the amount of SO_2 removed in the lower loops exceed the capacity of the oxidation air supply there. Module A remained out of service as a "spare."

Each module's upper-loop reaction tank was spiked to the desired DBA concentration, then each lower-loop reaction tank was spiked to the corresponding estimated steady-state value. Feed rates to each upper-loop reaction tank were set with the goal of maintaining a steady upper-loop DBA concentration. The system was allowed to approach steady state for 24 hours at full load. The first inventory was then conducted, marking the beginning of the test. The DBA

feed rates were adjusted twice a day to keep upper-loop reaction tank concentrations in the target range.

DBA inventories were conducted once each day. An inventory consisted of recording all tank levels, gauging the tanker trailer which was used as the on-site DBA storage vessel, recording liquor blowdown totalizer readings, and taking samples from each system vessel to analyze for concentrations of the DBA components.

2.3 Test Measurements

2.3.1 Flue Gas Sampling

During the baseline and parametric tests, the SO₂ removal performance measurements included the module inlet and outlet, and lower-loop outlet flue gas SO₂ concentrations. Inlet concentrations were measured using TECo's existing certified SO₂ analyzer. The accuracy of this analyzer was verified by pre-test EPA Reference Method 6 samples at the Module B inlet. Module outlet SO₂ concentrations were measured using Method 6 with a 24-point traverse across the outlet duct. SO₂ concentrations measured by TECo's on-line outlet monitor were also recorded for comparison, but only the Method 6 results were used in SO₂ removal calculations. Lower-loop outlet gas SO₂ concentrations were measured using an adaptation of EPA Method 6 with a flue gas sample pulled from a single test port on the module wall. The test port was situated beside the bowl collector that separates the two loops.

Prior to the baseline tests, the Module B outlet duct was traversed with a sample probe connected to a continuous SO₂ analyzer. This instrument indicated a significant variation in SO₂ concentration ranging across the outlet duct. This confirmed the need to traverse the duct to obtain a representative outlet concentration. Since TECo's on-line outlet SO₂ analyzer draws a gas sample from a single point in the duct, close agreement between the on-line meter and Method 6 results was not expected.

During a typical half-day test, the flue gas velocity was first measured at the scrubber module outlet by pitot tube traverse. Then, duplicate Method 6 traverses were done at the module outlet sample location, while duplicate single-point Method 6 samples were obtained at the lower-loop outlet sample location. For each pair of Method 6 samples at each location, a composite flue gas sample was obtained for Orsat analyses for CO₂ and O₂ concentration. For tests with longer durations, the half-day gas-sampling routine was repeated, so that four sets of lower-loop outlet and module outlet flue gas SO₂ measurements were usually obtained for each test day.

Flue gas SO₂ concentrations from the Method 6 samples were determined on site using the barium perchlorate titration procedure on aliquots from the Method 6 impinger solutions. These on-site analyses were used to verify that the results were reasonable and to make testing decisions. The remaining Method 6 impinger solutions were shipped to an off-site laboratory where the analyses were repeated using a more accurate ion chromatography (IC) method. The results from the IC method were used in all subsequent data reduction and only those results are reported here.

2.3.2 Slurry Sampling

During each baseline and parametric test, multiple sets of slurry samples were obtained from the upper and lower loops of the test module. These samples were collected concurrent with the Method 6 flue gas sampling efforts. Also, for each test, at least one liquor sample was collected from each loop. To collect the liquor samples, slurry from each loop was filtered at the sampling point and collected in prepared bottles of preserving solutions.

For every baseline test and each of the half-day parametric tests, three sets of slurry samples and one set of stabilized liquor samples were obtained. During one-day parametric tests, three sets of slurry samples and two sets of stabilized liquor samples were taken. For two-day parametric tests, the one-day schedule was doubled.

2.3.3 Chemical Analyses of Slurry and Filtrate Samples

For the baseline and parametric tests, all of the slurry samples were analyzed for suspended solids content and solid-phase carbonate content. These results were used to calculate limestone loading and utilization, which are important performance parameters. Limestone utilization can change relatively quickly with minor variations in operating pH.

Complete solid-phase analyses including calcium, magnesium, sulfite, and sulfate were done for only one of the half-day-test slurry samples and two of the one-day- and two-day-test slurry samples. The complete solid-phase analyses are used to calculate the oxidation fraction. This is also an important performance parameter, but the time constant for changes in slurry solids composition is much longer than for utilization. The oxidation fraction is also not as sensitive to minor changes in operating pH.

Settling rate tests and filter leaf tests were completed with slurry samples from the two-day baseline test and one- and two-day parametric tests. Settling tests were completed on site, while the filter leaf tests were conducted at an off-site laboratory. One solids sample from each of the two-day baseline and parametric tests was examined and photographed using scanning electron microscopy.

All of the stabilized liquor samples were analyzed for liquid-phase calcium, sulfite, sulfate, and carbonate. These results were used to estimate calcium sulfite, gypsum, and calcium carbonate relative saturations, which are important process chemistry indicators that can change relatively quickly with changes in test conditions, especially pH. Only one filtrate sample per day was analyzed for soluble species such as magnesium, sodium, and chloride. The time constant for changes in these soluble species concentrations is usually on the order of weeks. Therefore, these concentrations were not expected to vary significantly over the duration of each short-term test. A few liquid-phase samples from the baseline and parametric tests were also analyzed for 26 trace species using inductively coupled argon plasma emissions spectroscopy.

For the additive consumption tests, samples were collected to determine the FGD system DBA inventory and to determine DBA solution losses through the liquor blowdown stream. Slurry samples were gravimetrically analyzed for solids content. Slurry liquor and DBA tanker samples were analyzed for DBA components by ion exclusion chromatography (ICE). Concentrations of the three DBA component acids--succinic (C4), glutaric (C5), and adipic (C6) acids--were determined.

2.4 Baseline Test Conditions and Results

DBA also contains a small amount (up to 3%) of nitric acid. This strong acid is completely dissociated over the entire range of pH encountered in the limestone FGD process and, therefore, does not contribute to the buffering capacity of the DBA additive. The nitric acid component of DBA is neutralized by reaction with limestone to produce soluble nitrate ion in the process liquor. At typical DBA feed rates, the amount of reagent limestone consumed by this reaction is on the order of 0.01% of the total reagent consumption. Much more reagent is consumed and more nitrate is produced, for example, by absorption and neutralization of NO_2 from the flue gas. Because of the negligible contribution of nitric acid from DBA to the process material balance, the DBA supply was not analyzed for nitric acid.

2.4.1 SO₂ Removal Efficiency Results

Table 2-1 summarizes the test conditions and SO₂ removal efficiency results for the baseline tests. More detailed test results are presented in Appendix A.

The inlet SO₂ concentrations averaged about 2950 ppm (dry basis, corrected to 3% oxygen) during Test 1. Throughout the baseline test period, inlet SO₂ concentrations trended downward, averaging about 2650 ppm by the time Test 7 was completed.

Baseline Tests 1 through 4 were conducted at the normal full-load gas velocity through the test module, of approximately 7.5 ft/sec. For Baseline Test 1, the slurry pH set points

Table 2-1

Average Baseline Test Conditions and Results

Test No.	Lower Loop pH	Upper Loop pH	Flue Gas Velocity (ft/sec)	No. of Upper Loop Pumps Operating	SO ₂ Removal Efficiencies (%)	
					Lower Loop	Overall
1	4.1	5.8	7.5	2	16	93.6
2	3.5	6.2	8.4	2	13	96.5
3	5.2	6.2	7.4	2	7	97.3
4	2.8	5.0	7.9	2	*	78.2
5	3.9	5.6	11.2	2	0	84.5
6	5.3	6.1	9.4	3	6	96.0
7	5.1	5.9	10.8	2	17	89.5

* Mid-point SO₂ concentrations were greater than the inlet concentrations.

were maintained at approximately the normal values, averaging 4.1 for the lower loop and 5.8 for the upper loop. The overall SO₂ removal efficiency for the test module averaged 93.6%. The single-point lower-loop outlet gas SO₂ concentration measurements indicated that 16% SO₂ removal occurred in that portion of the module.

In Baseline Test 2, the upper-loop pH was increased to 6.2, which was estimated by Teco to be the maximum level that could be maintained without encountering sulfite blinding of the limestone reagent. Sulfite blinding can occur in a forced-oxidation system when the rate of SO₂ removal and/or when the liquor pH is high enough that the oxidation air flow rate is not sufficient to rapidly oxidize all of the removed SO₂. The lower-loop pH was allowed to equilibrate with no reagent addition, resulting in a pH of about 3.5. This low pH ensured high reagent utilization and complete oxidation of the lower-loop slurry. The higher operating pH in the upper loop increased the overall SO₂ removal to an average of 96.5%, which actually met the objective of 95 to 98% overall removal without DBA addition. With the lower operating pH in the lower loop, the SO₂ removal there decreased to about 13%.

Baseline Tests 3 and 4 were primarily conducted to collect FGDPRISM calibration data over the widest possible range of operating conditions. Baseline Test 3 was conducted at a higher pH set point of 5.2 in the lower loop and the same upper-loop set point of 6.2 as the previous test. The overall SO₂ removal efficiency for this test increased to an average of 97.3%. Although this is a high SO₂ removal efficiency, the lower limestone utilization in the lower-loop slurry at this pH would not allow the byproduct gypsum solids to meet wallboard specifications. Baseline Test 4 was conducted at low pH set points in the upper- and lower-loop slurries. With pH set points of 5.0 and 2.8, respectively, the overall SO₂ removal across the module dropped to about 78%.

Baseline Tests 5 through 7 were conducted at the highest achievable gas velocity through the module; the inlet vanes on the module booster fan were adjusted to the 100% open position. These tests were conducted to calibrate the model for conditions where the Unit 4 FGD system would be used to scrub flue gas from an adjacent unit.

For Baseline Test 5, the resulting flue gas velocity was about 11 ft/s. The pH set points were at the normal levels of 4.2 in the lower loop and 5.7 in the upper loop. The overall SO₂ removal efficiency averaged 84.5%, which is well below the value of 93.6% measured for Test 1 at the same pH but the normal gas velocity of 7.5 ft/s.

In Baseline Tests 6 and 7, the pH set points were increased to 5.1 in the lower loop and 6.1 in the upper loop. The booster fan inlet vanes remained 100% open. For Test 6, the spare upper-loop recycle pump was turned on, and for Test 7 this pump was again turned off. Operating the third absorber pump during Test 6 reduced the scrubber velocity to an average of 9.4 ft/s because of the increased pressure drop across the module. The flue gas velocity returned to nearly 11 ft/s when the third pump was turned off for Test 7.

Comparing results from Tests 5 and 7 at the same gas velocity and slurry flow rates shows that increasing the operating pH levels from normal to high increased the overall SO₂ removal efficiency from 84.5% to 89.5%. Comparing results of Tests 6 and 7 shows that activating the spare absorber pump increased the overall SO₂ removal from 89.5% to 96%, although some of this increase was due to the reduction in gas flow.

2.4.2 Baseline Test SO₂ Removal Performance Correlations

Absorber performance can be approximately described by the following expression:

$$\text{Number of Transfer Units (NTU)} = \ln (\text{SO}_{2\text{in}}/\text{SO}_{2\text{out}}) = K A/G \quad (2-1)$$

where: SO_{2in} and SO_{2out} = inlet and outlet SO₂ concentrations;
K (lb/hr-ft²) = average overall gas-phase mass transfer coefficient;
A (ft²) = total interfacial area for mass transfer; and
G (lb/hr) = total gas flow rate.

It is assumed in the above expression that the equilibrium partial pressure of SO₂ above the FGD liquor is small compared to the inlet and outlet concentrations.

The overall coefficient K can be expressed as a function of two individual coefficients, k_g and k_l , that represent mass transfer rates across the gas and liquid films, respectively:

$$1/K = 1/k_g + H/k_l\phi \quad (2-2)$$

where H is a Henry's law constant, and ϕ is the liquid-film "enhancement factor." For a given absorber operating at constant gas and liquid flow rates, NTU will be a function of slurry pH because of the effect of pH on the liquid-film enhancement factor. NTU will also be a function of DBA concentration for the same reason.

The form of Eq. 2-2 suggests that the effects of increasing pH (and DBA concentration in the later parametric tests) on the overall mass transfer coefficient, and therefore on NTU or SO₂ removal efficiency, will diminish at some point when H/k_l becomes small compared to $1/k_g$. This is referred to as "gas-film-limited" mass transfer. When this point is reached for a given absorber, there is no benefit to increasing the additive concentration.

Eq. 2-1 also shows that NTU should be inversely proportional to gas flow rate (if K and A are independent of gas velocity) and proportional to liquid flow rate (if A is proportional to liquid flow rate).

Figure 2-2 is a plot of NTU vs. pH for the lower-loop section of the absorber module during the baseline tests. In the figure, data for tests at three different gas velocities (7.5, 9.4, and 11 ft/s) are indicated by different symbols. Results for tests occurring at slightly different velocities have been normalized to these three velocities using Eq. 2-1.

It is evident from this plot that the scatter in the SO₂ removal data does not allow for any meaningful conclusions about the effects of pH and gas velocity on the lower-loop SO₂

removal performance. Two factors contribute to this scatter. First, the lower-loop exit flue gas samples are taken from a single point, and may not be representative of the entire gas stream. Second, because of the low removal efficiencies, the lower-loop exit flue gas concentration is not substantially different from the module inlet gas concentration. Consequently, normal sampling and analytical variabilities cause large errors in the calculated values for NTU.

The data in Figure 2-2 show that the lower-loop NTU did not exceed 0.25 and was generally less than 0.2. This is small compared to the overall NTU for the module (lower loop plus upper loop) which ranged from 1.5 to 3.8 for the baseline tests. Therefore, the lower-loop contribution to the overall mass transfer performance of the scrubber module was about 10% or less. Because of this, overall performance for the module can be approximately correlated with upper-loop recirculating slurry pH alone.

Figure 2-3 is a plot of overall module NTU vs. upper-loop slurry pH. In this plot, the data from Baseline Tests 1 through 4 at 7.5 ft/s, Tests 5 and 7 at 11 ft/s, and Test 6 at 9.4 ft/s (with three upper-loop slurry feed pumps) have been designated by different symbols so that the effects of gas velocity and slurry flow rate on NTU can be

illustrated. If K and A in Eq. 2-1 were independent of flue gas velocity, the NTU for tests at 11 ft/s would be about 70% of that for tests at 7.5 ft/s with the same slurry pH. The data in the figure are consistent with this expected effect. This suggests that increasing the gas velocity did not have a significant effect on the product of K and A .

The results for Test 6 at 9.4 ft/s with three upper-loop slurry pumps operating show that the performance increase due to increased slurry flow rate approximately offsets the expected 20% decrease in NTU due to increased gas velocity. Slurry flow rate measurements during the baseline tests showed that operating three pumps increased the spray rate to the absorber by a similar amount, ranging from 15% to 30%. The observed increase in NTU with slurry flow rate appears to be directly proportional to the slurry rate. This observation is consistent with data from EPRI's High Sulfur Test Center (HSTC) pilot unit.

Figure 2-2. Lower-Loop NTU vs. pH - Baseline Tests

Figure 2-3. Overall NTU vs. Upper-Loop pH - Baseline Tests

2.4.3 Results of Baseline Test Slurry Sample Chemical Analyses

Solids Analyses

The results of chemical analyses on slurry samples collected during the baseline tests are summarized in Appendix A. These analytical results are briefly described here.

The baseline test results in Appendix A show that limestone utilization in the lower loop is very high (99+%) at normal operating conditions. High utilization (>98%) is required to meet the wallboard-quality gypsum specification (<1.5% CaCO_3).

Limestone utilization is also high (92%) in the upper loop at normal operating conditions. High utilization in the upper loop is not required to maintain gypsum quality because only the lower-loop solids are purged to the dewatering system. Upper-loop utilization can affect the rate at which deposits accumulate in the packing and on the mist eliminator, however. From this standpoint, high utilization is also desirable in the upper loop.

Note that in Baseline Tests 2 and 3, where the upper-loop pH set point was raised to 6.2 to achieve high overall SO_2 removal levels, the upper-loop limestone utilization was measured at only 77 to 79%. These low upper-loop limestone utilization values suggest that high pH set points in the upper loop may not be a desirable approach for achieving 96 to 97% overall SO_2 removal with this FGD system.

Oxidation is calculated as $[1 - \text{moles of sulfite} / \text{moles of total S}]$. Sulfite was detected in only three of the 18 baseline solid samples analyzed. Under normal operating conditions in Test 1, oxidation was 100% in both the upper and lower loops. Oxidation remained at 100% in the upper loop under all test conditions, but sulfite was detected in the lower-loop samples in each of the high lower-loop pH tests (Baseline Tests 3, 6, and 7). As with high utilization, complete oxidation in the lower-loop is required to meet the gypsum specification (<0.5% CaSO_3). Higher pH set points have been observed to lower sulfite oxidation percentages in other

FGD systems. There are two likely effects of higher pH set points. One is that the amount of SO₂ removed increases relative to the amount of oxidation air which remains constant. The other is that liquid-phase bisulfite ion concentrations are reduced at higher pH values; bisulfite ion is thought to be a reactant in the sulfite oxidation reaction chain.

Liquid Analyses

As with the solids analyses, details of the baseline test liquid-phase analytical results are presented in Appendix A. The following paragraphs describe the trends observed in these results.

Gypsum relative saturations for the baseline test samples averaged about 1.06 in the lower loop and 1.12 in the upper loop, and did not appear to change significantly with changes in flue gas velocity or slurry pH. These relative saturation levels are well below the threshold value of 1.3 where gypsum scale formation might be expected to occur.

Calcium sulfite relative saturations were greater than 1.0 for only four samples, one each from Baseline Tests 3, 5, 6, and 7. Calcium sulfite relative saturations greater than 1.0 suggest the possibility of forming calcium sulfite solids in the FGD byproduct solids, and greatly increase the possibility of sulfite blinding of the limestone reagent. Solid sulfite was detected in slurry from the lower loops in three of these tests (3, 6, and 7), which is consistent with the higher calculated relative saturations of 2.9 to 6.0 in those samples. No solid sulfite was detected in the upper loop sample from Test 5. The calculated calcium sulfite relative saturation for this test was about 1.6, which apparently was not high enough to nucleate sulfite solids. These results are consistent Radian's previous experience, indicating that calcium sulfite relative saturations of approximately 3 or greater are required to nucleate sulfite solids.

Calcium carbonate relative saturations were highest in samples from the upper loop during high-pH tests (Baseline Tests 2, 3, 6, and 7). This is consistent with the observed limestone utilization values, which were lowest for these tests.

Chloride concentrations during the baseline tests averaged 4,300 ppm in the upper loop liquor and 14,800 ppm in the lower-loop liquor. The upper-loop chloride concentrations averaged 28% of the lower-loop values for the normal velocity tests and 33% for the high velocity tests. Chlorides enter the upper loop primarily by carryover from the lower loop, so these results indicate that carryover from the lower loop may have increased by about 20% at the higher gas velocity. The absorber materials of construction at Big Bend Unit 4 (C-276 alloy in the lower loop and 904L alloy in the upper loop) permit operation at much higher chloride levels than those measured during these tests. Therefore, the effect of higher flue gas velocity on chloride carryover from the lower to the upper loop is not expected to be an important factor in high-velocity operation at Big Bend.

Selected liquor samples from the baseline tests were also analyzed for trace metal species. These results are also included in Appendix A.

2.5 Parametric Test Conditions and Results

2.5.1 Parametric Test SO₂ Removal Efficiency

Table 2-2 summarizes the actual test conditions and SO₂ removal efficiency results for the parametric tests. More detailed results are tabulated in Appendix B.

During Parametric Test 1, the inlet SO₂ concentration measured by TECo's certified CEM averaged about 2530 ppm (dry basis, corrected to 3% oxygen). Inlet SO₂ concentrations ranged from about 2500 to 2900 during the three-week test period.

For all of the parametric tests, the lower-loop pH was held at 3.9 or lower, rather than the normal 4.2 set point. The reason for the lower set point was to avoid having the lower-loop SO₂ removal exceed the capacity of the oxidation air supply, to avoid the possibility of limestone blinding, and to ensure complete sulfite oxidation and high limestone utilization in that loop. These conditions are required for producing wallboard-quality gypsum.

Table 2-2

Average Parametric Test Conditions and Results

Test No.	Lower Loop pH	Upper Loop pH	Lower Loop DBA (ppm)	Upper Loop DBA (ppm)	Flue Gas Velocity (ft/sec)	No. of Upper Loop Pumps Operating	SO ₂ Removal Efficiencies (%)	
							Lower Loop	Overall
1	3.6	5.9	332	72	7.7	2	*	93.6
2	3.6	5.9	243	79	10.1	2	*	88.2
3	3.7	6.3	351	48	7.8	2	13	93.0
4	3.7	6.0	319	83	10.1	2	9	92.9
5	3.5	5.9	736	204	7.5	2	7	96.9
6	3.4	5.9	405	158	9.5	2	11	94.1
7	3.5	6.1	532	117	7.3	2	6	97.6
8	3.9	6.1	347	120	9.7	2	*	93.6
9	3.5	5.9	1143	291	7.4	2	8	97.9
10	3.5	5.9	1217	417	9.8	2	2	97.7
11	3.4	5.8	1023	441	9.2	3	11	99.0
12	3.7	6.1	1172	470	7.5	2	18	99.4
13	3.7	6.1	1180	423	9.8	2	10	98.2
14	3.6	6.1	2137	907	7.8	2	29	99.7
15	3.6	6.0	1800	881	9.9	2	13	99.1

* Mid-point SO₂ concentrations were greater than the inlet concentrations.

The results in the table show that the target SO₂ removal efficiencies for this program were readily achieved with DBA addition. At the normal flue gas velocity of 7.5 ft/s through the test module and a normal upper-loop pH, 98% overall SO₂ removal was achieved with an upper-loop DBA concentration between 250 and 300 ppm. The maximum SO₂ removal efficiency achieved was 99.7%, with a slightly elevated upper-loop pH and an upper-loop DBA concentration of about 900 ppm. At a higher flue gas velocity of nearly 10 ft/s, an upper-loop DBA concentration of approximately 400 ppm was required to achieve 98% removal, and the maximum SO₂ removal with 900 ppm of DBA was 99.1%. These results are discussed further in the next section.

As described earlier, during the parametric tests, DBA was added to both the upper- and lower-loop reaction tanks to simulate steady-state conditions. Under steady-state conditions, DBA would be added only to the upper loop and would reach the lower loop by way of slurry blowdown from the upper loop and liquor returned from the dewatering system. The data in Table 2-2 show that the ratios of DBA concentrations in the lower-loop versus upper-loop reaction tanks were widely varied, from a low of about 2:1 to a high of about 7:1, even though DBA feed rates were adjusted periodically in an attempt to maintain this ratio in the range of approximately 2:1. During the subsequent steady-state consumption test, DBA was added to the upper-loop tanks of all three operating modules and was not added to the lower-loop tanks. Under these conditions, the observed steady-state DBA concentration in the lower loop ranged from 1.6 to 2.4 times that in the upper loop, which confirmed the initial material balance estimates. Fortunately, the amount of SO₂ removed in the lower loop was only a small fraction of the total SO₂ removed in the test module, because the lower loop was operated at pH 3.9 or lower. Therefore, the variation in lower-loop DBA concentration during the parametric tests did not significantly affect overall SO₂ removal results. This is discussed further in the next section.

2.5.2 Parametric Test SO₂ Removal Performance Correlations

Figure 2-4 is a plot of NTU vs. DBA concentration for the lower-loop section of the dual-loop module during the parametric tests. In this figure, data for tests at different gas

Figure 2-4. Lower-Loop NTU vs. DBA Concentration - Parametric Tests

velocities are again indicated by different symbols. The plot shows that the scatter in the lower-loop SO_2 removal data seen in the baseline test results was also seen in the parametric test results.

The data show that the lower-loop NTU again did not exceed 0.2 for the parametric tests, except at the highest DBA concentration. This result is similar to the NTU values seen during the baseline tests. It shows that lowering the pH set point in the lower loop from the normal level of 4.2 in the baseline tests to 3.8 for the parametric tests was effective in preventing high SO_2 removal in the lower loop with the DBA concentrations tested. The effectiveness of DBA as a buffer is greatly reduced at pH levels less than 4 because much of the proton-accepting capacity of the DBA ions in solution is already depleted at this pH.

Because the lower-loop NTU for the parametric tests again represents, at most, 10% of the overall NTU for the module, overall performance can be approximately correlated with upper-loop slurry pH and DBA concentration. Figure 2-5 is a plot of overall module NTU vs. upper-loop DBA concentration at the normal and high levels of slurry pH for tests conducted at a scrubber gas velocity of 7.5 ft/s. Figure 2-6 is a similar plot for the tests at 10 ft/s gas velocity. In this plot, the data from Test 11 with three upper-loop slurry feed pumps operating have been designated by different symbols so that the effect of upper-loop slurry flow rate on NTU can be illustrated. Because the tests were not always conducted at these exact velocities, the results shown in the figures have been adjusted slightly to exactly 7.5 and 10 ft/s by assuming that the NTU is inversely proportional to the gas velocity (see Eq. 2-1).

In Figure 2-5, the overall NTU for the module operating at 7.5 ft/s is essentially linear with upper-loop DBA concentration from the lowest level to the mid-range concentration. The results for tests with a high upper-loop pH (6.1 to 6.2) lie about 0.5 to 0.7 NTU above those for the tests at normal pH (5.7 to 5.9). This effect of upper-loop pH on overall NTU is similar in magnitude to that observed for the baseline tests (see Figure 2-3). These trends indicate that, for this range of DBA concentrations and pH values, the mass transfer capability of the module has not yet approached the gas-film-limited level.

Figure 2-5. Overall NTU vs. Upper-Loop DBA Concentration (7.5 ft/sec)

Figure 2-6. Overall NTU vs. Upper-Loop DBA Concentration (10 ft/sec)

The results for Test 14, at the highest DBA level and high pH, lie significantly below the level that would be indicated by extrapolating the correlation from the lower DBA levels. This trend is expected based on the form of Eqs. 2-1 and 2-2. As the DBA concentration is increased to very high levels, the product of k_i and ϕ becomes large, and the overall mass transfer rate approaches the gas-film-limited rate (i.e., $1/K \approx 1/k_g$). The observation that this occurs at a very high SO_2 removal efficiency of 99.7% ($\text{NTU} = 6.4$) for the TECo scrubber module with only 900 ppm DBA in the upper loop indicates that this absorber is a good gas/liquid contactor. The spray tower absorber at EPRI's HSTC, for example, reached a maximum SO_2 removal efficiency of 96.6% ($\text{NTU} = 3.38$) at 5000 ppm DBA with an L/G comparable to that of the upper-loop section of the TECo scrubber.

In Figure 2-6, the results for the tests at high velocity (10 ft/s) show trends similar to those observed for the normal velocity tests. The difference in performance between the normal-pH and high-pH set points is less for the high velocity tests because a lower high-pH set point was used to avoid limestone blinding. Comparing NTU at the lower pH value for equivalent DBA levels, the increase in gas velocity from 7.5 ft/s (Figure 2-5) to 10 ft/s (Figure 2-6) decreased the NTU by about 20% to 25%. This result would be expected based on Eq. 2-1. This decrease in NTU with increasing velocity is similar in magnitude to what was seen in the baseline tests.

In Figure 2-6, the results for Test 11 with three upper-loop slurry pumps are only slightly higher than the results with two pumps operating. With three pumps operating, the slurry spray rate to the upper loop during Parametric Test 11 was measured to increase by only 6%. The observed increase in mass transfer capability was roughly proportional to the increase in slurry flow rate. This effect is consistent with the results from the baseline tests, and with data from EPRI's HSTC pilot unit. In all three cases, NTU was found to be directly proportional to slurry flow rate, for operation at constant pH and gas velocity.

2.5.3 Results of Parametric Test Slurry Sample Chemical Analyses

Solids Analyses

The results of chemical analyses on slurry samples collected during the parametric tests are summarized in Appendix B. These analytical results are briefly described here.

The results in Appendix B show that no significant change in the solids composition was observed during the parametric tests, compared to the baseline results. No sulfite was detected in the parametric test solids, primarily because no tests were conducted with a higher pH set point in the lower loop, as had been the case during the baseline test series. Limestone utilization was very high in all of the lower-loop samples from the parametric tests. This was also expected for operation at the lower-loop pH set point values of 3.5 to 3.9. Limestone utilization in the upper loop was similar to that observed for the baseline tests at comparable pH levels.

The lower-loop solids analyses for the parametric tests show that the carbonate and sulfite contents remained well below the upper limits for wallboard-grade gypsum. This would suggest that the lower-loop pH set point could be higher than the average of 3.8 selected for the DBA parametric tests. A minor improvement in SO₂ removal cost effectiveness might be obtained by raising the lower-loop set point, thus lowering the DBA concentration required to achieve a given SO₂ removal efficiency. However, results of liquid-phase analyses discussed below suggest that higher pH values should not be considered because of the potential for gypsum scale formation in the lower loop. If DBA were to be used on a continuing basis at Big Bend, the optimum lower-loop pH set point should be established with further testing.

Figure 2-7 compares the relationship between upper-loop limestone utilization and operating pH for the baseline and parametric tests. The data show that DBA

Figure 2-7. Upper-Loop Limestone Utilization vs. pH

addition had no significant effect on limestone utilization. The results in Figure 2-7 can be used as a guide to select a maximum feasible upper-loop operating pH. Based on previous experience, Radian recommends that limestone utilization be maintained above 85% to minimize scaling and plugging in packing or mist eliminators. Allowing for normal instrument and control variations, the data show that a pH of 5.9 would probably be the highest recommended set point.

Liquid Analyses

As with the solids analyses, details of the parametric test liquid-phase analytical results are presented in Appendix B. The following paragraphs describe the trends observed in these results.

Calculated gypsum relative saturations for the parametric liquid samples averaged 1.16 in the lower loop and 1.15 in the upper loop. The difference between the baseline and parametric averages for the lower-loop gypsum relative saturation is significant at the 99% confidence level, indicating that the lower loop operated at a higher relative saturation during the parametric tests (if there were no analytical bias differences between the test series). This result is unexpected, considering that the average lower-loop SO₂ removal efficiency was about the same for the baseline and parametric tests.

The highest calculated gypsum relative saturation values for the lower loop during the parametric tests were about 1.3, which is approaching the threshold level for gypsum scaling. These results indicate that gypsum scaling in the lower loop could possibly be a problem during operation with DBA addition. This possible problem could be avoided, however, by maintaining a lower pH set point in the lower-loop to reduce SO₂ removal and the amount of gypsum formed in that loop. Because these calculated gypsum relative saturations may not be accurate enough to draw firm conclusions regarding scaling potential, selection of an appropriate lower-loop pH set point with DBA addition should be based on further operating experience.

Calcium sulfite relative saturations for the parametric test liquid samples reached a maximum of about 1.6 in the upper loop during Test 13, which was conducted at high velocity and high pH. This relative saturation level is similar to that seen during the baseline tests under similar conditions. These results are again consistent with the solid-phase analyses which showed no calcium sulfite in the parametric test solid samples. As mentioned earlier in this section, previous experience suggests that a relative saturation of about 3 would be required to precipitate calcium sulfite in the solid phase.

As in the baseline test results, calcium carbonate relative saturations for the parametric test liquid samples were highest in the upper loop during tests at the high pH set point.

Chloride concentrations in the parametric tests averaged 3,200 ppm in the upper loop and 12,600 ppm in the lower loop for Tests 1, 3, and 5 at the normal flue gas velocity. The chloride concentration in the upper loop averaged about 25% of that in the lower loop, compared to 28% for the baseline tests at the same velocity. For tests at the high velocity (Tests 2, 4, and 6), chloride concentrations averaged 3,100 ppm in the upper loop, or about 32% of the lower-loop average of 11,800 ppm. This compares to 33% for the baseline tests at high velocity. Thus, the change in chloride distribution between the loops with increasing gas velocity appears to have been about the same for the baseline and parametric test series and, again, appears to be related to increased slurry carry-up at higher gas velocity.

In general, lower chloride (and sodium) concentrations were observed during the parametric tests than were measured during the baseline tests. This change was most likely a result of changing the system makeup water source to eliminate the use of a recycled pond water stream.

Chloride concentrations fell drastically during Parametric Tests 7 through 9 as a result of a malfunction in a mist eliminator wash valve on Module B. More than 650,000 gallons of excess makeup water were added to the system during this period. The analytical results in

Appendix B indicate that the normal system chemistry was re-established about one week later, before the long-term DBA consumption test began.

Selected liquor samples from the parametric tests were analyzed for trace metal species. These results are also included in Appendix B. The data show that most of the trace species concentrations were essentially unchanged for the two test series. There appear to be some increases in the concentrations of chromium, copper, and silicon, and a significant increase in iron, but these changes cannot be attributed to the use of DBA without knowing the normal variability of their concentrations in this system.

2.6 DBA Consumption Test Conditions and Results

The cost effectiveness of using additives to enhance SO₂ removal depends primarily upon the consumption rate of the additive. To provide a better measure of DBA consumption rates for this FGD system, a long-term DBA consumption test was performed on the entire Unit 4 FGD system. DBA was fed continuously to the upper loop of each of the three operating modules for nearly four days.

Following completion of the parametric tests, the steady-state, upper-loop DBA concentration necessary to obtain 98% SO₂ removal was determined from the parametric test data. Modules B, C, and D were operated with an upper-loop pH set point of approximately 5.6 but a reduced lower-loop pH set point of 3.8 to limit the amount of SO₂ removed in the lower loops with DBA present. Module A remained off line.

Each module's upper-loop reaction tank was spiked to a DBA concentration of approximately 350 ppm, and each lower-loop reaction tank was spiked to the estimated corresponding steady-state value of approximately 650 ppm. DBA feed rates to each upper-loop reaction tank were set with the goal of maintaining DBA concentrations there in the range of 300 to 400 ppm, based on estimates from a system material balance model and parametric test data.

The system was allowed to approach steady state for 24 hours at full load, then the first inventory was conducted, marking the beginning of the test.

The DBA feed rates were adjusted about twice a day to keep upper-loop reaction tank concentrations in the 300 to 400 ppm range. This was difficult, as the plant did not operate at full load during the overnight hours. The Unit 4 boiler operated at 90% to 100% load during the day and at 50% to 60% load overnight. In spite of this variation, the upper-loop concentration for the three modules averaged 321 ± 42 ppm DBA during the consumption test, with the range representing the 95% confidence interval.

As described in Section 2.2, DBA inventories were conducted once each day. Originally, seven inventories were to be taken. A boiler tube leak resulted in the shutdown of Unit 4, however, and only five inventories were actually conducted. The consumption test was to last 150 hours, but the boiler tube failure forced the end of the test at the 90-hour point. Samples from the first, fourth, and fifth inventories were analyzed for DBA content. The results of these analyses are included in Appendix C.

2.6.1 Consumption Rate Calculations

The following terms represent the mechanisms for loss of DBA from the FGD system during the consumption test:

- 1) Solution losses--primarily the DBA lost in blowdown liquor and in liquor adhering to the filter cake. Based on cake wash efficiency (95%), losses from the liquor entrained with filter cake solids were estimated to be a maximum of 0.3 lb/hr (<0.5% of addition rate), and were neglected in the consumption rate calculations. Incidental losses such as overflow splashed from the upper-loop reaction tanks were also neglected.
- 2) Degradation losses--losses which resulted from DBA components reacting to form byproducts other than adipic, glutaric, or succinic acid.

- 3) Coprecipitation losses--losses resulting from the incorporation of DBA into the calcium sulfate crystal structure. Based on experience at EPRI's HSTC, coprecipitation losses are most likely negligible for this forced-oxidation process.
- 4) Evaporative losses--vaporization of DBA from the system, mainly into the flue gas in the absorber. Also based on experience at EPRI's HSTC, these losses were considered to be negligible, as the vapor pressures of DBA components are very low at FGD process conditions.

The sum of losses by mechanisms 2, 3, and 4 is normally termed the "nonsolution loss."

To determine whether loss mechanism 3 (coprecipitation) would be significant for this FGD system, slurry samples from Parametric Test 15, at the highest DBA concentration tested in this FGD system, were also analyzed for DBA content in the solid phase. No DBA component was detected. However, the detection limit for the solid-phase analyses (0.4 mg/g) does not preclude the possibility that solid-phase DBA would be a significant term in the DBA material balance. Analyses for solid-phase DBA content under forced-oxidation conditions have also been conducted at EPRI's HSTC. At liquid-phase DBA concentrations up to 3250 ppm, no solid-phase DBA was detected in those samples either. For the Big Bend system's DBA material balance, the losses by mechanism 3 were therefore assumed to be insignificant. Since vaporization losses were also considered negligible at TECo's operating conditions, the nonsolution losses should result almost entirely from degradation.

The average nonsolution loss rate for a specific test period can be estimated using the following equation:

$$\text{Average Nonsolution Loss Rate Over Period } \Delta t = \frac{(\text{Addition})_{\text{Total}}}{\Delta t} - \frac{(\text{Blowdown})_{\text{Total}}}{\Delta t} - \frac{(I_t - I_0)}{\Delta t} \quad (2-3)$$

where, I represents the system DBA inventory.

The DBA addition term was obtained by measuring the change in the DBA tanker trailer level and the corresponding DBA solution analysis. The DBA blowdown term was obtained by multiplying the total blowdown (read from a totalizing meter) by the average DBA concentration of blowdown samples collected between totalizer readings. The inventory term was measured by a sampling and analysis effort on the major FGD system tanks and vessels, as described earlier.

The DBA nonsolution loss rate is normally reported on an SO₂ removal basis (lb of DBA per ton of SO₂ removed). The amount of SO₂ entering the FGD system was calculated using the unit load, unit heat rate, and daily coal analysis data. SO₂ emission rates were calculated using coal analyses, unit load, unit heat rate, and stack gas CEM SO₂ and O₂ concentration data.

2.6.2 Results

The results of the consumption rate computations are summarized below. Details of these results are tabulated in Appendix C.

DBA Loss Rates

Overall DBA nonsolution loss rates were calculated by summing the rates for the individual DBA components. The DBA loss rates are detailed in Appendix C. The total DBA loss rate was measured to average approximately 59 lb/hr, or 6.5 lb/ton of SO₂ removed. The nonsolution losses were measured to average 49 lb/hr, or 5.4 lb/ton of SO₂ removed. Of the total DBA lost from the system, 83% was through nonsolution loss mechanisms and 17% was through blowdown.

An error-propagation analysis was conducted to estimate the uncertainty in the nonsolution loss rate results. Uncertainty analyses were performed only for the total DBA nonsolution loss rates (in both lb/hr and lb/ton of SO₂ removed). The estimated errors for each measurement associated with the loss rates were input to an error-propagation analysis program

based on ANSI/ASME PTC 19.1-1985, "Measurement Uncertainty." On a lb/hr basis, the uncertainty is 14% of the mean; in lb/ton of SO₂ removed, the uncertainty is 19%. Therefore, the overall DBA nonsolution loss rate averages should be expressed as 49 ± 7 lb/hr, or 5.4 ± 1.0 lb/ton of SO₂ removed. Similar percentages can be assumed to hold for the individual component nonsolution loss rates.

SO₂ Removal Efficiency

One goal of the DBA consumption test was to achieve an average 98% SO₂ removal efficiency. The actual average removal efficiency was calculated to be short of that goal, at 95.3%. However, one of the three modules in service during this test, Module C, was operating at less than normal efficiency because a different, less efficient packing configuration was being used. It is very likely that the different packing configuration in Module C is entirely responsible for the removal efficiency averaging less than 98%.

During Baseline Test 1, for example, the Module C SO₂ removal efficiency was measured at about 80% compared to 94% for Module B. The target DBA concentrations required to achieve 98% removal efficiency were estimated from the parametric test results, which applied to the more efficient Module B. Based on the comparison data from Baseline Test 1, if Module B was in fact operating at a 98% SO₂ removal efficiency, we estimate that the Module C efficiency would have been increased to about 89% or 90% from the baseline level of 80%. The overall performance for two modules operating at 98% removal efficiency and one module operating at 90% efficiency would be very near the observed consumption test average of 95.3%.

DBA Degradation Products

Although only succinic, glutaric, and adipic acid concentrations were measured and are reported in Appendix C, it is possible that some degradation products which also contribute to the slurry liquor buffer capacity can be formed. Examination of ICE chromatographs from all samples revealed significant concentrations of at least three components that are most likely

degradation products. Since no degradation product standards were available, the chemical structure of these components is unknown. Two components were eluted prior to succinic acid (C4) and one eluted between succinic and glutaric (C5).

Buffer capacity titrations performed during the parametric and long-term tests over-predicted the DBA concentrations (as measured by ICE) by 10% to 20% in the modules and up to 50% in the dewatering area. This discrepancy could potentially be explained by the generation of degradation products which contribute to the liquor buffer capacity. However, no further analyses were performed to attempt to identify the degradation products to determine if they would, in fact, be expected to contribute to the liquor buffering capacity.

DBA Consumption in the Waste Handling Area

At the conclusion of the parametric tests, TECo operating personnel changed the sludge surge tank vessel in service in the waste handling area. These vessels are used to store the underflow from the primary dewatering hydroclones until it can be processed in the secondary dewatering vacuum filters. Surge tank "A," which contained a high DBA concentration following the final parametric test, was taken out of service (not processed) during the consumption test, offering the opportunity to determine if DBA degrades during the long slurry residence times often encountered in the waste-handling portion of the FGD system (particularly in a system that has a thickener for primary dewatering). Previous research has shown DBA degradation to occur primarily in the presence of oxidation reactions (such as sulfite oxidation) that occur in the FGD system absorber, although some biological degradation has been observed by Radian to occur in liquor held in ponds.

In addition, DBA concentrations in the active waste-handling vessels were examined to determine if DBA levels changed dramatically. These active vessels included a blowdown slurry surge tank (upstream of the hydroclones), the active sludge surge tank "B" (downstream of the hydroclones), a filtrate storage tank (downstream of the vacuum filters), and a return water surge tank (downstream of the filtrate storage tank).

The results reveal an interesting phenomenon. After the "A" sludge storage tank (downstream of the hydroclones) was idle for 40 hours, analyses showed a marked change in the distribution of the DBA components. In 115 hours time, both the adipic acid and succinic acid concentrations had fallen below the detection limit. However, glutaric acid concentrations did not change significantly over the entire period of observation (130 hours).

In the active blowdown tank, which the first slurry handling tank in the dewatering system, DBA components had similar concentrations as in the modules. From that tank, slurry goes to the hydroclones for primary dewatering. Hydroclone underflow then entered sludge surge tank "B", the feed vessel to the vacuum filtration unit. Recall that sludge surge tank "A" was idle during this time period. In sludge surge tank "B", the adipic acid concentration averaged 65% lower than in the active blowdown tank; glutaric and succinic acids averaged 28% lower and 82% lower, respectively. This indicates that some form of degradation was occurring in the active sludge surge tank "B", possibly similar to that seen in the idle vessel, sludge surge tank "A".

The mechanism for the observed degradation of succinic and adipic acids in the dewatering area has not been determined, but a biological degradation mechanism seems possible. It is interesting to note that glutaric acid, which appeared to be relatively unaffected by whatever mechanism degraded adipic and succinic acids, has the median molecular weight of the three.

2.7 Effect of DBA Addition on Solids Properties

During all three test periods (baseline, parametric, and DBA consumption), tests were performed to examine the effect of DBA on solids properties. If DBA is used as a performance-enhancing additive, changes in solids properties caused directly or indirectly by DBA addition could affect the operation of dewatering equipment.

Three methods were used to determine solids properties during this test program: settling tests, filter leaf tests, and scanning electron microscopy (SEM). Settling tests were performed on site on lower-loop slurry samples to document any effect of DBA addition on

gypsum byproduct solids settling rates and final settled densities. Filter leaf tests were performed to assess changes in the solids filtration rate and water retention after vacuum filtration. Finally, SEM was used to qualitatively examine changes in crystal structures and sizes.

2.7.1 Settling Tests

Batch settling tests were performed on slurry from the Module B lower loop to determine both settling rates and final solids underflow concentrations. The bulk settling procedure detailed in Method C2 of EPRI's FGD Chemistry and Analytical Methods Handbook was followed (however, rake action was not simulated). Settling rates are reported as the unit area (in ft²/ton/day) required to reach a 30 wt.% underflow concentration. Detailed results are included in Appendix D.

The Module B slurry settling rate and final underflow solids concentration changed markedly during the parametric tests. Figure 2-8 plots the unit areas and final underflow data against upper-loop DBA concentration. The data are plotted versus upper-loop DBA concentration rather than lower-loop concentration because it is the upper-loop concentration that is controlled to enhance system SO₂ removal performance.

In Figure 2-8, the unit areas increased from 0.8 ft²/ton/day for the baseline samples to 4.1 ft²/ton/day at 440 ppm DBA in the upper loop. This seems to indicate that the settling rate slowed with increasing absorber DBA levels. The solids also compacted less at the higher DBA levels. The final underflow concentration decreased from 65 wt.% solids for the baseline samples to 45 wt.% at 440 ppm DBA in the upper loop.

Also, at the beginning of the long-term consumption test, the lower-loop solids were observed to transfer from the "hindered-settling" regime into the "particulate settling" regime. That is, rather than settling as a bulk mass, most of the solids fell freely to the cylinder bottom. Some fine particulates settled in a hindered manner. This phenomenon was observed in each of the three settling tests performed during the DBA consumption test. These results are not

Figure 2-8. Settling Test Results

reported in Appendix D, as the settling test could not be performed properly under "particle settling" conditions. These settling test data suggest that there is a correlation between upper-loop DBA concentration and the solids settling rate. There are two possible explanations for this correlation: 1) DBA or other factors could have affected solids properties directly or indirectly by altering other process conditions; or 2) other independent process variations such as system fines content could have affected the test results.

The operating data and chemical analyses results from this site, as well as results from other sites where DBA has been added to FGD systems, provide no support for the first explanation. DBA has not been reported to adversely affect solids dewatering properties at other sites. Furthermore, there were no major changes in other solids properties upon DBA addition that might have affected dewatering properties. For example, sulfite oxidation, limestone utilization, or solids residence time in the FGD system were not significantly affected.

A review of TECo's past operating experiences suggests that the second possible explanation is more likely the cause of the change in solids settling properties during the DBA addition periods. TECo has experienced problems with fines buildup in the system, as fines are not effectively removed by the hydroclone primary dewatering devices. A buildup of fines can alter slurry particle size distributions. It is possible that a buildup of fines content in the Module B lower loop during the parametric test period caused the trend of slower settling. No particle size distribution data were taken to further investigate this possibility, however.

2.7.2 Filter Leaf Tests

Filter leaf tests simulate the performance of a rotary drum vacuum filter. Two separate tests were performed: form filtration and cake moisture. The form filtration test was performed on lower-loop slurry samples to determine the effective solids filtration rates (lb/hr/ft² filter area). The test was performed by measuring the time from the start of the filtration apparatus until the first cracks appeared on the surface of the filter cake. The test results give an indication of the required filtration surface and indicate the ease with which water is drawn from

the solids. Samples were adjusted initially to 30 wt.% solids so that individual test results could be compared. The cake moisture test was performed by applying a vacuum to a sample for a constant time period and measuring the water content in the resulting cake. This test measures the tendency of the filtered solids to retain water.

Filter leaf test samples were taken concurrently with settling test samples. The filter leaf test conditions and results are also included in Appendix D. Filtration rates varied from approximately 300 to 1000 lb/hr/ft²; cake solids contents ranged from 77 wt.% to 86 wt.%. The data are plotted as a function of upper-loop DBA concentration in Figure 2-9. Note that there is no apparent correlation between DBA concentration and either filtration rate or cake solids content.

In addition to filter leaf samples from the Module B lower loop, TECo vacuum filter operating data were examined to determine the effect of DBA on dewatering properties. During the baseline tests, the filter product solids content averaged 89.4 wt.%. During the parametric tests, the product solids content fell steadily from 85.3 wt.% to 83.2 wt.%. During the DBA consumption rate test, the vacuum filter product solids content averaged higher than during the parametric tests, at 86.9%. TECo's current gypsum specifications require a minimum of 87% solids. These operating data suggest that the addition of DBA resulted in a reduction in gypsum product solids content.

However, it should be noted that TECo has previously experienced occasional difficulties in meeting the gypsum moisture content specification, with no DBA in the system. In those previous instances, the vacuum filter cake moisture content was determined to have been adversely affected by a buildup of fines in the system, as described above. Blowdown of fines-containing liquor from the return water tank has been employed to mitigate this problem. The effect of fines on the filter cake moisture content may well have confounded any effect of the DBA additive on solids properties.

Figure 2-9. Filter Leaf Test Results

2.7.3 SEM Analyses

SEM was used to detect changes in crystal morphology. SEM photographs of solids from Baseline Test 1, Parametric Tests 1, 5, and 9, and the consumption rate test (first inventory) are shown in Figures 2-10a through 2-10e. As with the filter leaf samples, SEM samples were taken concurrently with settling test samples. Since sample preparation can change the particle size distribution, the SEM pictures should only be used in a qualitative evaluation. Crystals from the parametric tests generally appear to have suffered less from attrition. Crystals from Parametric Tests 1 and 5 seem to have a more plate-like habit than those from Baseline Test 1 or Parametric Test 9. No correlation between crystal morphology and DBA concentration could be established, however. These results suggest that it is unlikely that changes in crystal morphology caused the slower settling of slurry solids during the parametric tests.

The SEM photograph from the long-term test did not differ significantly from the parametric results, offering no explanation for the free-settling phenomenon observed during settling tests for this period.

2.7.4 Wallboard Gypsum Evaluation

A sample of gypsum byproduct produced at Big Bend during the DBA consumption test was submitted by TECo to Gold Bond Building Products, a division of National Gypsum. Gold Bond is currently producing wallboard from TECo's gypsum byproduct. Gold Bond reported that no DBA was detectable in the sample. This is indicative that the filter cake washing in the vacuum filters at Big Bend is very effective at removing DBA as well as chlorides in the liquor adhered to the dewatered solids. Gold Bond also reported that all physical and chemical tests yielded acceptable results within the range of properties measured by gypsum produced without DBA additive in the FGD system.

a. Baseline Test 1

b. Parametric Test 1

Figure 2-10. SEMs of Samples from TECo Big Bend Baseline and Parametric Testing

c. Parametric Test 5

d. Parametric Test 9

Figure 2-10. Continued

e. Consumption Test Inventory 1

Figure 2-10. Continued

Thus, in spite of the possible adverse effects of DBA addition on gypsum dewatering properties that were observed during the parametric and DBA consumption tests, the end user of the wallboard-grade gypsum byproduct did not note any adverse effects. This further supports the contention that the observed increase in solids settling rate and decrease in filter cake solids content were not a direct result of DBA addition, but instead a result of some confounding effect such as a gradual buildup of fines in the FGD system during this test period.

3.0 FGDPRISM SIMULATIONS

The EPRI FGD Process Integration and Simulation Model (FGDPRISM) is a computer program that simulates the performance of FGD systems. The model was calibrated to TECo's Big Bend Station Unit 4 Module B scrubber using data obtained during four of the baseline tests and one of the parametric tests. After calibration, the model results were compared to results from other baseline and parametric tests. The model was found to adequately predict the performance observed in most of the tests not used for the calibration. Then, the calibrated model was used to investigate the effects of DBA, packing height, and L/G on SO₂ removal. Each of these efforts is described in a separate subsection below.

3.1 FGDPRISM Calibration

The latest version of the FGDPRISM model (Version 2.0) was calibrated to the TECo Big Bend 4 system. Modifications included in this latest version enabled more accurate modeling of packing in the absorber, and more rigorous gas/liquid interface calculations than could the previous version (Version 1.1). The Version 2.0 dual-loop template was also modified to allow for two more spray headers in the lower loop (previous versions only allowed for one spray header) since the TECo Big Bend 4 FGD system has three headers in the lower-loop section.

The upper-loop section of the absorber was divided into three separate modules for modeling purposes--the headers above the packing, the packing, and the spray section below the packing. The majority of the gas/liquid mass transfer within the Big Bend 4 FGD system occurs in the upper loop, primarily in the packing. Therefore, the interfacial area of the packing is an important parameter in calculating the overall system SO₂ removal. The surface area was calculated directly from the specific area of the packing (42 ft²/ft³ for Big Bend 4) and the tower volume occupied by the packing.

During the calibration, more emphasis was placed on matching the upper-loop performance because more of the removal occurs in that loop, especially in the cases where there was a low pH in the lower loop. Details about Version 2.0 of FGDPRISM and about the calibration of the model to the Big Bend Unit 4 FGD system are discussed in Appendix E.

Table 3-1 compares the actual SO₂ removal, pH, limestone utilization, and oxidation values measured during the five tests used for the calibration to the values predicted by FGDPRISM. Figures 3-1 and 3-2 illustrate how well the calibration fit the measured values for overall and upper-loop SO₂ removal, respectively.

These results show that the calibrated model predicted overall and upper-loop SO₂ removal well. Agreement was not as good, however, for the lower-loop SO₂ removal efficiency. This did not affect the overall results significantly because most of the SO₂ removal occurs in the upper loop. The calibration procedure for the dual-loop process simulation required that a large number of cases be run because there is a separate set of calibration parameters for the lower and upper loops. The effort required to further improve the lower-loop results was not considered cost-effective considering the minimal impact on predicting overall SO₂ removal efficiency.

The data in Table 3-1 show that the calibrated model did well at predicting pH values for both loops. For the test cases with very little excess limestone in the lower loop (Baseline Tests 2 and 5, and Parametric Test 9-Run 6), the Version 2.0 calibration matched the lower-loop pH values better than in previous calibrations. The earlier version of FGDPRISM had difficulties with these low reagent ratio cases. The measured lower-loop pH values for these cases ranged from 3.45 to 4.03, and the pH values predicted with Version 2.0 ranged from 3.88 to 4.11.

For the upper-loop pH values, the largest difference between the measured and predicted pH was for Baseline Test 5, with pH 5.63 measured and pH 5.91 predicted by the model. However, referring to the plot of upper-loop limestone utilization versus upper-loop

Table 3-1

Comparison of Observed and Predicted Performances for the Calibration Cases

			Baseline Test 1		Baseline Test 2		Baseline Test 3	
			Observed	Predicted	Observed	Predicted	Observed	Predicted
Removal (%) -	Overall		93.6	93.4	96.4	96.5	97.3	98.5
	Upper Loop		92.3	92.8	95.9	96.3	97.1	98.3
	Lower Loop		16.0	7.9	12.5	6.4	7.0	12.9
Overall NTU			2.75	2.72	3.32	3.35	3.61	4.20
Slurry pH -	Upper Loop		5.7-5.8	5.91	6.2-6.3	6.07	6.2-6.3	6.08
	Lower Loop		4.1-4.3	3.98	3.4-3.9	4.02	5.1-5.2	4.89
Utilization (%) -	Upper Loop		91.9	92.1	77.6	77.4	79.1	78.8
	Lower Loop		99.1	98.7	99.1	98.9	80.3	80.3
Oxidation (%) -	Upper Loop		100.0	99.0*	100.0	99.0*	100.0	98.0*
	Lower Loop		100.0	99.5*	100.0	99.0*	90.2	93.0*

			Baseline Test 5		Parametric Test 9, Run 6	
			Observed	Predicted	Observed	Predicted
Removal (%) -	Overall		84.5	84.2	98.9	98.7
	Upper Loop		83.8	83.5	98.6	98.4
	Lower Loop		4.0	4.7	17.0	15.6
Overall NTU			1.86	1.85	4.61	4.34
Slurry pH -	Upper Loop		5.6-5.7	5.91	5.9	5.96
	Lower Loop		3.9-4.2	4.10	3.4-3.5	3.87
Utilization (%) -	Upper Loop		88.2	88.0	89.6	89.6
	Lower Loop		98.9	98.8	99.8	99.4
Oxidation (%) -	Upper Loop		100.0	99.0*	100.0	99.3*
	Lower Loop		100.0	99.0*	100.0	99.5*

* Percent oxidation is not predicted by the model; these are the oxidation percentages that were input to the model to match the observed liquid-phase sulfite concentrations.

Figure 3-1. Comparison of FGDPRIISM Predictions and Observed Values for Overall SO₂ Removal for the Calibration Cases

Figure 3-2. Comparison of FGDPRIISM Predictions and Observed Values for Upper-Loop SO₂ Removal for the Calibration Cases

slurry pH in Figure 2-7, the data point for Baseline Test 5 seems to be an outlier. As can be seen in that figure, the higher pH predicted by FGDPRISM is more consistent with the other pH values that were observed to correspond with a similar utilization value.

3.2 Model Fit to Other Test Results

Results from only one of the parametric tests and four of the seven baseline tests were used to calibrate FGDPRISM for this FGD system. Results from the other baseline and parametric tests can be compared to model predictions to verify the calibrated FGDPRISM's ability to model the effects of DBA and other process variables on the SO₂ removal performance of the Unit 4 FGD system.

The best way to determine the model fit to these other test data is to run the model for the actual conditions of each test, then compare model predictions to the measured SO₂ removal data. This involves varying the limestone reagent ratio; concentrations of soluble species such as magnesium, sodium, and chloride; DBA concentration; and other parameters to match the observed conditions for each test.

Five parametric tests, covering the range of conditions evaluated in the parametric test series, and two baseline tests were selected for such an evaluation. Table 3-2 summarizes the conditions for the tests selected for comparison and key results of the FGDPRISM simulations for these cases. The model predictions for SO₂ removal performance are compared to the test results in Figure 3-3.

The model predicted the observed SO₂ removal performance reasonably well for the seven comparison tests. The observed SO₂ removals for these seven tests ranged from 84.5 to 99.7%. For the tests at the lower end of this range, the predicted performance was greater than the observed performance by 1.5 to 3 percentage points. However, the model predictions at these lower SO₂ removal percentages are of little interest for this program.

Table 3-2

Comparison of Observed and Predicted Performance for Noncalibration Cases

	Parametric Test 2 Run 3		Parametric Test 3 Run 2		Parametric Test 5 Run 1	
	Observed	Predicted	Observed	Predicted	Observed	Predicted
Removal, %						
Overall	88.1	91.3	96.8	97.3	96.0	95.9
Absorber	88.6	90.9	96.4	97.2	96.1	95.6
Quench	--	4.0	10.9	6.0	--	7.4
Overall NTU	2.17	2.44	3.44	3.61	3.22	3.19
Slurry pH						
Absorber	5.91	5.91	6.19	6.14	5.94	5.93
Quench	3.59	3.75	3.45	3.87	3.54	3.72
Utilization, %						
Absorber	93.2	93.4	80.6	81.3	93.8	94.0
Quench	99.8	99.7	99.7	99.5	99.9	99.6
Absorber Chemistry						
Cl	3112	3184	2781	2880	3436	3450
Mg	1479	1485	1362	1351	1555	1531
Ca	930	861	927	871	934	962
SO ₃	107	89	86	64	65	80
SO ₄	4680	4573	4507	4208	4207	4173
DBA	69	69	48	54	214	261
Quench Chemistry						
Cl	10973	8802	11967	9551	15958	13318
Mg	3489	3662	3571	3815	4718	4627
Ca	1633	1082	1679	1183	1868	1585
SO ₃	128	1703	95	1606	132	1490
SO ₄	5332	6938	4440	6466	4448	5537
DBA	149	165	351	150	1155	750

Table 3-2

Continued

	Parametric Test 11 Run 2		Parametric Test 14 Run 2		Baseline Test 5		Baseline Test 7	
	Observed	Predicted	Observed	Predicted	Observed	Predicted	Observed	Predicted
Removal, %								
Overall	99.0	96.8	99.7	98.9	84.5	86.2	89.5	91.0
Absorber	98.9	96.5	99.6	98.6	83.8	85.5	87.4	90.1
Quench	10.9	9.2	23.3	17.3	4.0	4.7	16.5	9.2
Overall NTU	4.61	3.44	5.81	4.51	1.86	1.98	2.25	2.41
Slurry pH								
Absorber	5.80	5.87	5.95	5.97	5.63	5.87	5.91	5.95
Quench	3.67	3.99	3.60	3.95	4.03	4.11	5.11	4.82
Utilization, %								
Absorber	94.2	94.3	86.9	86.8	88.2	93.2	87.8	92.1
Quench	99.7	99.3	99.9	99.2	98.9	99.3	86.4	85.9
Absorber Chemistry								
Cl	3029	3099	3294	3315	4998	4746	3900	4058
Mg	1679	1742	1477	1504	1621	1678	1280	1259
Ca	943	844	1002	1029	1200	1003	975	1020
SO ₃	145	106	35	74	212	86	38	69
SO ₄	5236	5283	3982	4096	3862	4414	3545	3713
DBA	441	494	907	905	0	0	0	0
Quench Chemistry								
Cl	9245	7427	10929	9542	16059	14093	13896	10936
Mg	3507	3803	3678	3791	4471	4378	2541	3047
Ca	1469	918	1386	1189	2251	1439	4269	1418
SO ₃	143	1699	98	837	234	1174	652	517
SO ₄	7234	8323	5270	6371	3775	6072	4054	4932
DBA	1023	1023	2137	2200	0	0	0	0

Figure 3-3. Comparison of FGDPRIISM Predictions and Observed Values for Overall SO₂ Removal for Baseline and Parametric Tests Not Used as Calibration Cases

For two tests nearer the middle of the SO₂ removal range (96 to 97%), the model predictions were very close, within 0.5 percentage points of the observed values. This is the range of greatest interest for this high-efficiency SO₂ removal project.

Finally, for the two tests where very high SO₂ removal had been observed (99.0% and 99.7%), the model tended to underpredict the observed values by one to two percentage points. On an NTU basis, the disagreement at these high removal levels was more significant, with the model underpredicting the observed performance by 1.2 to 1.3 NTU. For the economic analyses of upgrade options, the accuracy of the SO₂ removal predictions on a percent removal basis are more important than as NTU values. The errors associated with the underprediction for these two tests (about 1 to 2 percentage points) are relatively minor. This, combined with the fact that this performance level is outside the range of most interest for this program (95 to 98% overall SO₂ removal), means that underpredicting for these two tests should not raise a serious concern about the use of model predictions to evaluate the upgrade options. Also, considering the potential difficulties associated with measuring very high SO₂ removal efficiencies across full-scale FGD absorbers, it is possible that the discrepancies between the model predictions and the observed removals are at least partially a result of measurement uncertainties.

3.3 Predictive Simulations

The calibrated Version 2.0 model was used to study the effects of adding DBA to the system and to predict the effects of other potential SO₂ removal upgrade approaches. These modeled approaches included operating at a different L/G (varying the number of upper-loop slurry recirculation pumps in operation) and removing or adding packing in the absorber, both in combination with DBA addition. The results of these predictive simulations were used in the economic analysis of the various options for increasing SO₂ removal, which is discussed in Section 4.

A base case was set up to simulate typical operating conditions for the Unit 4 FGD system. The inputs used were similar to those for Baseline Test 1, with a few exceptions. The

upper-loop limestone utilization was increased to 93.5%, and the inlet SO₂ concentration was decreased to 2735 ppm (dry basis), as these values were more representative of other baseline test conditions. FGDPRISM predicted an overall SO₂ removal of 92.1% for this base case with 2 feet of packing. The upper-loop removal was predicted to be 91.5% with a pH of 5.87, and the lower-loop removal was 7.5% with a pH of 3.96. The results of Baseline Test 1, which is used as the base case for the economics in Section 4, showed a similar average overall SO₂ removal of 93.3%.

Packing heights of one and three feet were also modeled. Lower-loop DBA concentrations of 0, 500, 1000, 1500, 2000, and 2500 ppm were then simulated for all three configurations. These lower-loop DBA concentrations correspond to upper-loop concentrations of approximately 0, 180, 400, 650, 900, and 1200 ppm, respectively. Because of the water balance configuration at Big Bend Unit 4, slurry at the lower-loop DBA concentration is fed to the balance of the system. Although for the Big Bend system the upper-loop DBA concentration would be controlled to vary the SO₂ removal performance, with FGDPRISM the system-wide DBA concentration is actually the model input. The corresponding upper-loop concentration is then predicted by the model.

In all of these simulations, the input reagent ratio was held constant at the base case conditions. At a constant reagent ratio, the model predicts that the upper-loop pH would drop slightly (about 0.1 pH unit) as the DBA concentrations are increased to the highest levels.

Figure 3-4 illustrates the predicted effects of increasing upper-loop DBA concentration and varied packing height. For the baseline case with two feet of packing, overall SO₂ removal was predicted to increase from 92.1% to 98.9% with the addition of 1163 ppm DBA in the upper loop (2500 ppm in the lower loop). This corresponds to a 78% increase in overall NTU from 2.54 to 4.51. With only one foot of packing in the absorber, overall removal increased from 76.1% with no DBA to 92.9% with 1176 ppm DBA in the upper loop. For this case, the overall NTU increased from 1.43 to 2.64, an increase of 85%. The effect of DBA weakens with three

Figure 3-4. Predicted Overall NTU vs. Upper Loop DBA Concentration with Gas Velocity = 7.5 ft/s

feet of packing in the absorber because the baseline removal is already high. The SO₂ removal with three feet of packing and no DBA is 98.0%. Adding 1181 ppm to the upper loop increased the removal to 99.8% in the upper loop and overall. The corresponding NTU increase was from 3.91 to 6.21, or 59%.

This model prediction shows that a higher DBA concentration is required to achieve 99% overall SO₂ removal than would be predicted from the parametric test data plotted in Figure 2-5. Actually, there are no data plotted in that figure for operation at a normal upper-loop pH and at upper-loop DBA concentrations above approximately 400 ppm. However, a minor extrapolation of those data suggests that less than 500 ppm of DBA would be required in the upper loop to achieve 99% overall SO₂ removal. As mentioned above, the calibrated model predicts that over 1100 ppm would be required.

There are several possible explanations for this apparent discrepancy. One is that, as discussed previously in this section, the current FGDPRIISM calibration underpredicts the observed SO₂ removal performance for very high removal levels (99% or greater). Another is that the model predictions illustrated in Figure 3-4 were run at slightly lower reagent ratio values than were experienced in the parametric tests, and thus higher DBA concentrations would be required to achieve equivalent SO₂ removal performance. Finally, also as discussed earlier, the potential measurement uncertainties associated with evaluating very high SO₂ removal across full-scale FGD modules may contribute to this discrepancy.

These apparent discrepancies will be addressed in the subsequent economic evaluations in two manners. First, for the configuration actually tested (two feet of packing and two slurry recirculation pumps in service in the upper loop), the actual parametric test data will be interpolated to predict the relationship between upper-loop DBA concentration and overall SO₂ removal. For the other cases not tested at full scale (varied packing level and/or varied number of pumps in service), the model predictions will be used. However, the effect of DBA concentration on the cost effectiveness of those options will be evaluated in a cost sensitivity analysis.

Figure 3-4 illustrates that FGDPRISM predicts a significant effect of packing height on SO₂ removal. Reducing the amount of packing from two feet to one foot dropped the overall SO₂ removal from 92.1% to 76.1% for the base case (no DBA). Adding another foot of packing increased the base case removal to 98.0%. The overall NTUs predicted were 1.43, 2.54, and 3.91 for one, two, and three feet of packing, respectively. The model predicted a larger increase (1.37 NTU) for adding one foot of packing than the decrease predicted (1.11 NTU) for removing one foot of packing. A linear relationship between the packing surface area and NTU would be expected if only the packing area was changing. This is not the case, however, as the overall mass transfer coefficient is predicted to increase as a result of increased limestone dissolution across the greater amount of packing.

Although increasing the packing height to three feet appears to be advantageous, the ability of the model to predict the effect of the packing area on SO₂ removal performance has not been validated for this system. If this appeared to be a technically and economically attractive option, then additional testing would be recommended to confirm the modeled effects of packing height.

One set of cases was run at a reduced upper-loop slurry recirculation rate of 18,000 gpm (60% of normal). This was to predict the effect of turning off one of the upper-loop recycle pumps. DBA was added to determine the amount required to achieve removals similar to normal operation. With 416 ppm DBA in the upper loop (1000 ppm in the lower loop), the overall removal was 91.5%, which is slightly less than the baseline removal. To achieve approximately 99% removal, 2300 ppm of DBA was required in the upper loop (4500 ppm in the lower loop). This simulation yielded 98.9% overall removal and 98.7% upper-loop removal. At the normal L/G, the model predicted that it would take 1163 ppm DBA in the upper loop to reach approximately the same removals. The low L/G simulation results are also included in Figure 3-4.

The economic analysis in the following section uses the results of these predictive simulations to weigh the benefits and costs of changing the L/G in the upper loop, adding DBA, or changing the amount of packing. Combinations of these options are also considered.

4.0 ECONOMIC EVALUATION

Performance data from the baseline, parametric, and long-term DBA consumption tests have been used to evaluate the economics of using DBA additive to increase SO₂ removal efficiency with the Big Bend Unit 4 FGD system. FGDPRISM predictions have also been used to look at the economics of other possible upgrade options that were not actually tested. The approach taken for the economic evaluations and the results of these evaluations are presented and discussed in this section.

4.1 Options Considered and Cost Basis

Table 4-1 summarizes the upgrade options evaluated for Big Bend Unit 4 and lists the cost components that were considered in the evaluation of each option. The first option is to add DBA to the system while maintaining the current operating conditions in the upper loop (a slurry pH of 5.7 to 5.8, two-ft packing depth, and two slurry recirculation pumps operating). The performance data obtained during the parametric and DBA consumption tests were used directly to evaluate this option. The DBA concentration and consumption rate required to achieve a given level of SO₂ removal efficiency were interpolated from the data reported in Section 2.

The installed capital costs of a DBA additive system for Big Bend Unit 4 were estimated by Stone & Webster Engineering Corporation (SWEC). Budgetary quotes for delivered cost of the DBA additive were provided by two suppliers: Monsanto and DuPont. Other costs for the first upgrade option in Table 4-1 include additional limestone reagent, operating and maintenance charges for increased gypsum production, and offsetting revenues for increased gypsum sales. Most of these cost data have been provided by TECo.

The other upgrade options listed in Table 4-1 have been evaluated on the basis of FGDPRISM predictions, using results from the calibrated model. There are no

Table 4-1

Upgrade Options for Big Bend Unit 4 FGD System

Option	Cost Components	Comments
Add DBA using current system configuration.	DBA Capital DBA Additive Additional O&M Additional Limestone Additional Gypsum Revenue	Can be evaluated using test data.
Add or remove packing (with DBA additive).	Same as Option 1 plus Packing Capital Fan Power	Requires FGDPRIISM simulation. Requires fan power versus packing depth data.
Change number of upper loop slurry recycle pumps operating (with DBA additive).	Same as Option 1 plus Pump Power Fan Power Nozzle Capital	Requires FGDPRIISM simulation. Requires fan power versus L/G data.

full-scale performance results available to verify these predictions. However, later in this section, a cost sensitivity analysis shows the effects of varying the concentration of DBA required to achieve a given level of SO_2 removal on the upgrade economics for these options. This sensitivity analysis should address the uncertainty associated with using modeled results for the economic analyses of these options.

The second upgrade option listed in Table 4-1 is to change the packing depth in the modules. DBA could also be used with this option to achieve higher SO_2 removal levels. To evaluate the economics of this upgrade approach, FGDPRISM simulations were used to estimate SO_2 removal efficiencies for either 1-ft or 3-ft packing depths at various DBA concentrations.

Packing capital costs for evaluating this option were provided by TECo. It is assumed that packing would be replaced annually as is the current practice. Changes in packing depth will also change the module pressure drop and fan power consumption. No full-scale data are available for fan power requirements as a function of packing depth. Consequently, values had to be estimated for the changes in module pressure drop and fan power consumption with the variation in packing depth. Later in this section, a sensitivity analysis shows the effect of varying this estimate on the upgrade economics.

The third upgrade option in the table is to change the number of operating upper-loop slurry recycle pumps (i.e., varying the upper-loop L/G). Again, DBA could be used with this option. Performance data were obtained during baseline and parametric testing with three rather than two upper-loop slurry recycle pumps operating. These data suggest that the overall NTU will be directly proportional to L/G. FGDPRISM has also been used to evaluate this option and has predicted a similar relationship between absorber L/G and NTU.

Some other equipment changes would be required to change the number of operating absorber recycle pumps. With only one pump operating, it is likely that one of the spray headers would need to be blanked off, or the nozzle sizes on the spray headers reduced, to keep the pump motor within its design operating range and to keep the spray nozzles operating at

the proper slurry feed pressures. With three pumps operating, it is likely that larger nozzles would need to be installed on the spray headers to take better advantage of the additional pump capacity. With the existing nozzles, much of the additional pumping capacity with three pumps in service goes into higher nozzle pressure rather than to increasing the L/G.

Data for module pressure drop (and fan power) versus L/G are also necessary for an accurate evaluation. To evaluate this option, the module pressure drop has been estimated as a function of L/G from baseline test data with three versus two pumps operating. The sensitivity of the upgrade economics to this variation in pressure drop with L/G is also addressed later in the section.

Table 4-2 summarizes the values used for the various economic factors mentioned above. The development of these values is discussed further in Appendix F.

4.2 Results

4.2.1 Economics for DBA Addition in the Current FGD Configuration

The economic factors described above were included in spreadsheet calculations to estimate the costs for additional tons of SO₂ removed by the various SO₂ removal efficiency upgrade options. Table 4-3 summarizes the cases for DBA addition in the current operating configuration (two feet of packing and two operating pumps in the upper-loop section) based on a delivered price of \$0.21/lb for DBA and a 65% Unit 4 capacity factor.

The SO₂ removal performance values in Table 4-3 are based on the actual parametric test performance correlation, which relates overall NTU to absorber DBA concentration for two feet of packing, with two upper-loop slurry recycle pumps operating, and with the normal pH set points. This correlation was illustrated in Figure 2-5. The expected overall SO₂ removal efficiency is entered in Column 3 for increasing levels of DBA concentration, which are listed in Column 2.

Table 4-2

**Economic Factors Used to Compare SO₂ Removal Upgrade Options
for Big Bend Unit 4**

Cost Item	Cost or Range	Source	Comments
DBA Capital	\$550,000 for 100 lb/hr system; cost vs. capacity adjusted using (capacity/100) ^{0.15}	SWEC	This cost is about 50% higher than that used in EPRI's FGDCOST spreadsheet.
DBA Additive	\$0.21 to 0.33/lb delivered for pure DBA shipped as a 50% solution	SWEC per Monsanto and DuPont	
Additional O&M	\$1.30/ton of gypsum	TECo	
Additional Reagent	\$7.80/ton of prepared limestone	TECo	
Gypsum Revenue	\$2.50/ton	Assumed	
Packing Capital	\$100,000 1st foot \$ 84,000 2nd foot \$ 68,000 3rd foot	TECo	Total cost for four modules (packing is replaced annually).
Packing Pressure Drop	0.5, 1.0 inch H ₂ O per foot	TECo	Range of ΔP's evaluated.
Fan Power @ \$0.05/kWh	\$66,000/yr per inch of added pressure drop (total for three modules)	Calculated based on 65% capacity factor assuming 70% fan efficiency	Actual cost could be somewhat lower because fan efficiency should increase as inlet vanes are opened.
Pump Power @ \$0.05/kWh	\$345,000/yr for one pump x three modules	Calculated from test data (287 kW per pump)	L/G for one pump was assumed to be 60% of that for two pumps, assuming one header is blocked off or nozzle sizes are changed; L/G for three pumps was assumed to be 140% of that for two pumps, assuming that the nozzle size is changed.
Module Pressure Drop Changes with L/G	±2 inches H ₂ O for adding or turning off pump, from normal 2-pump operation	Estimated from test data	
Nozzle Capital	\$96,000	Calculated assuming \$200 per nozzle for 480 nozzles	

Table 4-3

Base Case Economic Calculations
\$0.21/lb of DBA, 65% Capacity Factor, 1" ΔP/ft of Packing

Packing Depth (ft)	Upper Loop DBA Conc. (ppm)	SO ₂ Removal (%)	Additional Tons of SO ₂ Available (per year)	Change in Total Annual Cost (\$1000/yr)	Marginal Cost (\$/ton SO ₂)	Average Cost (\$/ton SO ₂)	Net Annual Value (\$1000)	
							@ \$250/ton	@ \$150/ton
Base Case								
2	0	93.3	5051	0	0	0	0	0
Add DBA								
2	125	95.5	3386	130	78	78	287	120
2	250	97.5	1858	190	39	59	608	289
2	500	99.2	619	287	78	65	822	378
2	750	99.6	307	371	271	78	815	341
2	1000	99.8	169	452	585	93	769	281

Column 4 shows the tons of SO₂ still available for removal on an annual basis. This number is based on there being 13.2 tons/hr of SO₂ in the Unit 4 FGD system inlet gas at full load, the SO₂ removal efficiency shown in Column 3, and a 65% Unit 4 capacity factor.

Column 5 shows the total net annual cost of each case, compared to the base case. As described earlier, these costs consist of annualized capital costs for an additive feed system, the delivered cost of DBA used, the delivered cost of additional limestone, the costs of grinding and feeding the limestone slurry, and the credit associated with producing additional gypsum. These individual cost components are detailed for each case in Appendix F, Table F-1. Example calculations are also included in that appendix.

Column 6 is the calculated marginal cost of additional tons of SO₂ removed, in dollars per additional ton removed, as the DBA concentration is increased from case to case. For the first of the DBA addition cases, the results in this column show that the lowest level of DBA concentration (125 ppm in the upper loop) yields additional SO₂ removed at a cost of \$78/ton. For the next increment, as the DBA concentration is increased from 125 to 250 ppm, additional tons of SO₂ removed are less expensive (\$39/ton). This is primarily because the capital cost of the DBA system is not very sensitive to DBA feed rate. The first DBA case includes the bulk of the capital investment, while the increased costs for the subsequent cases reflect only the costs associated with enlarging the feed system. Again, example calculations are given in Appendix F.

Column 7 also reflects the cost of additional tons of SO₂ removed as the DBA concentration is increased but calculated in a different manner. The values in Column 6 are referred to the case immediately above in the table, while the values in Column 7 are all referred back to the base case. Thus, the values in Column 7 represent an average marginal cost for all of the additional tons of SO₂ removed when going from baseline performance to the removal efficiency listed in Column 3.

The final two columns show calculated net annual values for additional tons of SO₂ removed. This is defined as the annual value of the additional tons removed minus the additional

annual costs, as were summarized in Column 5. The results in Column 8 assume that emissions credits are valued at \$250/ton, while the results in Column 9 assume they are valued at a more conservative value of only \$150/ton.

By examining the marginal cost of additional tons of SO₂ removed in Column 6, we see that the costs rise sharply as the absorber approaches its gas-film-limited performance. The marginal cost associated with raising the upper-loop DBA concentration from 500 ppm to 750 ppm, correspondingly raising the overall SO₂ removal performance from 99.2% to 99.6%, is estimated at \$271 per additional ton of SO₂ removed. This is greater than even the larger of the two SO₂ allowance values considered, indicating that it would not be cost effective to raise the upper-loop DBA concentration to 750 ppm to achieve this higher overall SO₂ removal level. The prior increment in the table, going from 250 ppm to 500 ppm DBA in the upper loop, and from 97.5% to 99.2% overall SO₂ removal, is much more cost effective. The marginal cost for this increment is only \$78/ton.

Correspondingly, the net annual value of the DBA additive upgrade peaks at about 99.2% SO₂ removal (500 ppm DBA in the upper loop) at about \$820,000 per year based on the higher \$250/ton allowance value. Even with a more conservative value of \$150/ton for SO₂ allowances, this case is still very cost effective, with a net annual value of about \$380,000. At this performance level, approximately 4400 additional tons of SO₂ could be removed by the Unit 4 FGD system (compared to baseline performance) at an average marginal cost of \$65/ton.

4.2.2 Economics for Modeled Upgrades

Two other upgrade options were considered: varying the depth of packing in the upper-loop sections of the modules and changing the number of operating pumps. These potential upgrade options were not tested at full scale. SO₂ removal performance results for these options are based on predictions from the calibrated FGDPRIISM model (see Section 3).

The costs for these options were also evaluated in spreadsheet calculations. Table 4-4 summarizes the cost estimates for these cases. Also included in the table is the base case performance. This base case is used as a reference point for the marginal costs (Column 6) in the first case for each option, and as the reference point for all of the "average cost" calculations (Column 7). Details of the costs are summarized in Table 4-4 and example calculations are presented in Appendix F.

The first upgrade option evaluated in Table 4-4 is adding 1 foot of packing to the upper-loop sections of the modules. FGDPRISM predicts a large increase in removal efficiency (from 93.3% with 2 feet of packing to 98.0% with 3 feet of packing), without DBA addition. Even for this no-DBA case, some costs are associated with adding the additional foot of packing. The capital cost for the third foot of packing is estimated at a total of \$68,000 per year for the four modules. The additional flue gas pressure drop (1 in. H₂O per foot of packing) increases fan power costs by about \$66,000/yr.

The marginal cost of the additional SO₂ removal (compared to the base case) achieved by adding packing alone is about \$47/ton. The estimated net annual value of this upgrade is about \$720,000 for a \$250/ton SO₂ allowance value, which is about \$100,000 less than for the optimum DBA-only case discussed previously.

The subsequent entries in the table show that adding DBA to this case with 3 feet of packing may or may not be cost effective, depending on the value of the additional SO₂ credits. The marginal cost of the first increment of SO₂ removal above 98% is \$204/ton, which is within the range of the two SO₂ credit values considered in Table 4-4. At a \$250/ton credit value, adding DBA to this option would be cost effective and would raise the net annual value to almost \$770,000 per year at an upper-loop DBA concentration of 250 ppm and 99.2% overall SO₂ removal. However, at a \$150/ton value, adding DBA would not be cost effective.

Table 4-4

Modeled Case Economic Calculations
\$0.21/lb of DBA, 65% Capacity Factor, 1 in. H₂O/ft of Packing

Packing Depth	Upper Loop DBA Conc., ppm	SO ₂ Removal	Addtl Tons SO ₂ Available	Change in Total Annual Cost, \$1000/yr	Marginal Cost, \$/ton SO ₂	Average Cost, \$/ton SO ₂	Net Annual Value (\$1000)		
								@ \$250/ton	@ \$150/ton
Base Case (2 ft of packing, 2 upper-loop pumps in service, no DBA)									
2	0	93.3	5051	0	0	0	0	0	0
With 3 ft of packing									
3	0	98.0	1506	167	47	47	719	365	365
3	125	98.8	914	288	204	70	747	333	333
3	250	99.2	637	336	176	76	767	326	326
3	500	99.5	357	423	311	90	750	281	281
3	750	99.7	196	506	514	104	708	222	222
3	1000	99.8	151	586	1783	120	639	149	149
With 1 ft of packing									
1	0	76.1	17987	-270	21*	21*	-4122	-1671	-1671
1	125	79.2	15637	-141	55	13*	-3663	-1447	-1447
1	250	81.9	13594	-81	29	10*	-3212	-1200	-1200
1	500	86.5	10172	29	32	-6*	-2467	-798	-798
1	750	89.7	7765	131	42	-48*	-1966	-538	-538
1	1000	91.9	6108	225	57	-213*	-1646	-383	-383

Table 4-4

(Continued)

Packing Depth	Upper Loop DBA Conc., ppm	SO ₂ Removal	Addtl Tons SO ₂ Available	Change in Total Annual Cost, \$1000/yr	Marginal Cost, \$/ton SO ₂	Average Cost, \$/ton SO ₂	Net Annual Value (\$1000)	
							@ \$250/ton	@ \$150/ton
With one absorber pump (plus nozzle change); 60% of flow @ 50% of power								
2	0	80.2	14874	-552	58 ^a	58 ^a	-1888	-906
2	250	89.2	8082	-335	32	111 ^a	-406	-119
2	500	92.4	5695	-231	44	359 ^a	86	135
2	1000	96.1	2914	-45	67	-21	595	365
2	2000	98.7	1000	288	174	71	741	320
2	2500	99.2	606	448	405	101	679	219
With 3 pumps (plus nozzle change); 140% of flow @ 150% of power								
2	0	97.7	1715	524	152	157	326	-24
2	250	99.4	423	698	134	151	475	-3
2	500	99.9	91	785	264	158	471	-41
2	1000	99.98	15	947	2124	188	328	-191

*These cases are referred back to the base case, but are predicted to achieve less than base case SO₂ removal. The \$/ton value listed represents a cost savings per ton of SO₂ not removed, compared to the base case.

The next series of cases in Table 4-4 is based on removing 1 foot of packing. The first calculated marginal cost for this series of cases is actually a "savings," as the net cost is reduced, but the tons of SO₂ removed are also reduced. The annual value of this case is negative, though, since the value of the credits at even \$150/ton far exceeds the \$21/ton saved. This option is clearly not attractive. The savings in annual packing cost and in fan power costs are far outweighed by the loss of performance. Adding DBA does not restore even baseline SO₂ removal performance with as much as 1000 ppm in the upper loop. At this condition, the total annual cost is estimated to increase by nearly \$225,000, but the overall SO₂ removal is actually lower than for baseline conditions.

The next upgrade option is based on taking one of the two operating upper-loop slurry recycle pumps out of service, with the normal two feet of packing in the modules. FGDPRISM predicts that NTU is roughly proportional to the upper-loop slurry flow rate, which is consistent with the limited L/G test results obtained by operating the third pump. For this case, it was assumed that 60% of the normal flow can be obtained with one pump because the frictional losses in the slurry piping will decrease. It was also assumed that the upper-loop spray nozzles would be replaced to restore the nozzle pressures at the reduced slurry rate, but that no other equipment changes would be required. The total annual costs in Table 4-4 includes pump power cost savings and an annual capital charge for the nozzle replacement.

Operation with one slurry recycle pump results in substantial savings in pump power (\$345,000/yr) and fan power (\$132,000/yr) relative to the base configuration. However, this option requires considerably higher DBA concentrations to achieve high-efficiency SO₂ removal than for the base configuration cases in Table 4-3. For example, 2500 ppm of DBA is predicted to be required in the upper loop to achieve 99.2% overall SO₂ removal, whereas this same level could be achieved with two feet of packing and two upper-loop pumps operating at only 500 ppm in the upper loop. Furthermore, this option is only attractive at moderate and higher DBA concentrations. Below 500 ppm DBA in the upper loop, the predicted SO₂ removal is well below the baseline removal, and the savings in operating cost do not offset the potential value of the SO₂ credits lost.

The optimum net annual value is achieved with this one-pump option at a 2000 ppm DBA concentration in the upper loop, at 98.7% overall SO₂ removal. For the \$250/ton SO₂ credit value, the optimum net annual value for this option is about \$80,000/yr less than the optimum for DBA addition in the base configuration, and about \$25,000/yr less than for the increased-packing option. At the lower \$150/ton SO₂ credit value, there is little difference between the optimum net annual values for these three upgrade options. The technical feasibility of this one-pump option is uncertain, however, as scaling in the upper-loop packing and plugging in the slurry piping are both possible consequences of operating with only one pump in service.

The final option evaluated in Table 4-4 is operation with three absorber slurry recycle pumps. This option is clearly less attractive than DBA addition in the base configuration, adding a third foot of packing or operation with one pump. At the higher, \$250/ton SO₂ credit value, the maximum net annual value is only about half that of the other options. At the lower \$150/ton value, the net annual values are all negative.

4.3 Cost Sensitivity Analyses

4.3.1 Effect of DBA Cost

The results presented above are based on an estimated price of \$0.21/lb for DBA. This is probably the lowest price that could actually be negotiated. To determine whether a higher DBA price would affect the relative attractiveness of the options, the calculations were repeated with a DBA price of \$0.33/lb.

At \$0.33/lb, the marginal cost of the first additional tons removed by adding DBA in the base configuration increases from about \$78/ton to \$90/ton. The maximum annual value of the DBA-only option is reduced to about \$740,000, based on the \$250/ton SO₂ credit value, and the added-packing case becomes nearly as attractive. At the lower, \$150/ton SO₂ credit value, the added-packing case actually becomes more cost-effective than DBA addition alone. The optimum net annual value for the case with 3 feet of packing and no DBA addition is estimated at

\$365,000, while the optimum for DBA addition in the current configuration is reduced by the higher DBA price to just under \$300,000 per year.

At the higher DBA cost, one-pump operation at higher DBA concentrations becomes clearly less favorable than either the DBA-only or added-packing cases, regardless of the SO₂ credit value considered.

4.3.2 Effect of DBA Concentration Requirement

For the modeled cases in Table 4-4, it is possible that the actual DBA concentration required to achieve a given SO₂ removal level will be lower than what FGDPRISM has predicted. This may particularly be true for the cases where very high overall SO₂ removal efficiencies have been predicted (e.g., 99% removal or greater). It was determined in the previous section that the current FGDPRISM calibration tends to underpredict the observed performance at very high SO₂ removal levels.

For example, the parametric test data suggested that 99% SO₂ removal could be achieved with approximately 500 ppm of DBA in the upper loop, whereas the current calibration of FGDPRISM predicted that over 1100 ppm would be required. In Table 4-3, where the economics of DBA addition with two feet of packing and two recirculation pumps in service are calculated, the actual parametric test data were used to estimate the relationship between DBA concentration and SO₂ removal performance. For the options not tested at full scale, in Table 4-4, the model-predicted relationship was used.

This suggests that for some of the modeled high SO₂ removal cases in Table 4-4, the DBA requirements may be overstated; the actual DBA requirements to achieve the stated level of SO₂ removal performance could prove to be no more than 50% of the predicted values. To assess the impacts of potential model overprediction of DBA requirements to achieve very high SO₂ removal efficiencies, the cases for three feet of packing and for one recirculating slurry pump operating were repeated with only half the original DBA concentration requirements.

In the cases with three feet of packing, the system SO₂ removal prediction was not highly dependent on the effects of DBA. If the concentration of DBA required to achieve 99% or greater overall SO₂ removal proved to be only half of what was predicted with the current FGDPRISM calibration, the maximum annual value of this option would be increased slightly, making it essentially equal in value to the base DBA addition (two feet of packing) case. At a credit value of \$150/ton of SO₂, the maximum net annual value for three feet of packing would be raised to about \$370,000/yr, while at the \$250/ton credit value, the maximum net annual value would be about \$830,000/yr.

For the cases with only one upper-loop slurry recirculation pump in service, the FGD system performance is largely dependent on DBA addition to achieve high SO₂ removal efficiencies. If the actual DBA concentration required to achieve a given level of SO₂ removal proved to be only half of what the current FGDPRISM calibration predicted, it would have a significant effect on the economics. At a credit value of \$250 for additional tons of SO₂ removed, the net annual value for this option increases to over \$1,000,000, making this potentially the most attractive option considered. Similarly, for a \$150/ton SO₂ credit value the net annual value estimate increases to over \$600,000, which again would make this potentially the most cost-effective option.

4.3.3 Effect of Packing Pressure Drop

If the actual packing pressure drop is less than the value assumed for Table 4-4 (1 in. H₂O per foot), the annual value of the added-packing option increases slightly. For example, at 0.5 in. H₂O per foot and with a \$250/ton SO₂ credit value and a \$0.21/lb DBA cost, the optimum annual value increases from \$770,000 to \$800,000. When DBA is evaluated at a cost of \$0.33/lb, the increase in value due to the lower pressure drop assumption is sufficient to make this option slightly favorable to the DBA-only option.

4.3.4 Effect of L/G on Module Pressure Drop

The effect of varying the number of operating upper-loop slurry recirculation pumps on module flue gas pressure drop was evaluated in Table 4-4 at +2 in. H₂O for three-pump operation and -2 in. H₂O for one-pump operation. This effect was estimated from a limited amount of data, so the economics for the one- pump and three-pump options were also calculated with values of ± 1 in. H₂O and ± 3 in. H₂O. The cost effectiveness of three-pump operation was not altered over this range of values; this option remains unattractive because of the increased pump power costs. For the one-pump case, the higher (3 in. H₂O) credit raised the optimum annual value of this case to within \$15,000 of the DBA-only option at a \$250/ton SO₂ credit value and with a \$0.21/lb DBA price. However, the attractiveness of this option remains very sensitive to DBA price, because of the high DBA concentrations required to maintain high-efficiency SO₂ removal operation. At the higher DBA cost of \$0.33/lb, even the 3 in. H₂O credit leaves this option clearly less cost effective than either DBA alone or adding packing. These numbers are all based on the model-predicted DBA concentrations for this configuration.

4.4 Recommended Upgrade Options

The economic evaluations described above indicate that it would be very cost effective to upgrade the SO₂ removal performance of the Big Bend Unit 4 FGD system to generate additional SO₂ allowance credits. There are three potentially favorable upgrade options: using DBA additive in the current module configuration, increasing the packing depth in the upper loop with or without DBA additive, or operating at higher DBA concentrations but with only one slurry recirculation pump in service in the upper loop. The DBA-only option was slightly more cost effective than the other two options under most conditions. Exceptions were when the highest potential DBA price of \$0.33/lb was considered, or when the DBA required by the two alternates was evaluated at half the model-predicted concentration.

In the optimum case for the DBA-only option, an additional 4400 tons of SO₂ could be removed at a net cost of only \$65 per additional ton removed. This estimated cost is

well below anticipated market prices for SO₂ allowances, and also well below the costs for adding SO₂ removal capability to an unscrubbed unit.

The DBA-only option is the recommended approach for achieving high-efficiency SO₂ removal operation with the Big Bend Unit 4 FGD system. There are several reasons for selecting this option over increased packing in the upper-loop section or operating at higher DBA concentrations but with only one upper-loop slurry recirculation pump in service:

First, evaluation of the DBA-only option was based on actual test data. The economic optimum DBA concentration appears to be about 500 ppm in the upper loop, which is somewhat higher than was used in the long-term consumption test, but the SO₂ removal performance and DBA consumption rate can be extrapolated to this higher concentration with reasonable confidence. Evaluations of increased packing depth and operating with only one pump, on the other hand, were based solely on FGDPRISM performance predictions. At this time, there are no full-scale data available to confirm the FGDPRISM predictions for these configurations. Therefore, there is more uncertainty in the economics of these options.

Other issues have not been addressed in the economic evaluation of the increased packing option. These are the effect of packing depth on plugging and/or scaling tendencies, and on the ability to clean the packing in place. Increased packing depths tend to increase recirculating slurry hold-up in the upper loop. This can increase the amount of limestone which dissolves while the slurry is within the absorber module, and can, in turn, lead to higher gypsum relative saturations and localized scaling in the packed section. Furthermore, TECo is trying to develop an approach for cleaning the packing with a high-pressure wash so that the practice of discarding plugged packing can be eliminated. Increased packing depth would make in-place cleaning much more difficult. For these reasons, TECo does not feel that increasing packing depth to upgrade SO₂ removal performance is an attractive option.

The third potentially attractive upgrade option, of operating with only one upper-loop slurry recirculation pump in service but at higher DBA concentrations, could also result in

operating problems that were not addressed in this economic evaluation. For example, localized gypsum scaling in the upper-loop packed section may result because of the increased amount of SO_2 removed per volume of slurry recirculated, and the slurry recirculation headers may plug because reduced velocities in the headers may not keep all of the slurry solids entrained. Therefore, this potential upgrade option would not be recommended for further consideration unless Tampa Electric decides to test this approach at full scale. Both the DBA concentration required to achieve high SO_2 removal efficiency and the tendency for scaling or plugging in the packing and slurry headers should be evaluated.

APPENDIX A

Results of Chemical Analyses for Baseline Tests

Table A-1

Actual Baseline Test Conditions and Results

Test No.	Run No.	Date (1992)	Lower Loop pH ^a	Upper Loop pH ^a	Flue Gas Velocity (ft/s)	No. of Upper Loop Pumps Operating	Orsat O ₂ (%)	Lear Siegler Inlet SO ₂ ^b (ppmv)	M6 Lower Loop SO ₂ ^b (ppmv)	M6 Outlet SO ₂ ^b (ppmv)	DuPont Outlet SO ₂ ^b (ppmv)	SO ₂ Removal Efficiencies (%)	
												Lower Loop	Overall
1	1	9-29	3.7/ 4.2	5.3/ 5.8	7.5	2	7.7	2934	2570	193	174	12	93.4
1	2	9-29	4.1/4.2/4.3	5.7/5.8/5.8		2		2973	2366	191	156	20	93.6
1	3	9-29	4.1/4.1/4.2	5.8/5.8/5.8	7.5	2		2947	2464	186	153	16	93.7
2	1	9-30	3.5/3.6/3.9	6.2/6.2/6.2	8.4	2	8.1	2734	2406	103	108	12	96.2
2	2	9-30	3.4/3.6/3.9	6.2/6.2/6.3		2		2759	2413	92	105	13	96.7
3	1	9-30	5.2/5.2/5.1	6.2/6.2/6.3		2	8.3	2735	2433	71	70	11	97.4
3	2	9-30	5.1/5.2/5.1	6.2/6.2/6.3	7.4	2		2690	2614	75	75	3	97.2
4	1	10-1	2.9/2.7/3.2	5.0/5.1/5.0	7.9	2	6.6	2735	2788	557	476	^d	79.6
4	2	10-1	2.8/2.8/3.2	5.0/5.0/4.9		2		2724	2807	635	487	^d	76.7
5	1	10-1	3.9/4.0/4.2	5.6/5.7/5.6	11.1	2	6.5	2759	3248	416	337	^d	84.9
5	2	10-1	3.9/4.0/4.2	5.6/5.7/5.6	11.2	2		2771	2757	441	335	0	84.1
6	1	10-2	5.3/5.3/5.1	6.1/6.0/6.0	9.7	3	6.5	2732	2641	107	112	3	96.1
6	2	10-2	5.3/5.3/5.1	6.0/6.0/6.0	9.1	3		2731	2519	113	108	8	95.9
7	1	10-2	5.2/5.3/5.1	6.0/6.1/6.0	10.8	2	7.1	2661	2224	240	204	16	91.0
7	2	10-2	4.9/5.3/5.1	5.8/6.0/6.0	10.8	2		2633	2193	315	229	17	88.0

^aReported pH values are Radian measurements, local on-line indicator, and control room data, respectively.

^bAll SO₂ concentrations are on a dry gas basis, corrected to 3% oxygen.

^cRemoval efficiencies based on off-site M6 analyses.

^dMid-point SO₂ concentrations are greater than the inlet concentrations.

Table A-2

Chemical Analyses Conducted for Baseline Tests

	Number of Analyses (Each Loop)	Technique
TEST 1		
Liquid-Phase Analyses		
Calcium	3	Atomic Absorption Spectroscopy (AAS)
Magnesium	2	AAS
Sodium	2	AAS
Chloride	2	Ion Chromatography (IC)
Sulfite	3	Iodometric Titration
Sulfate	3	IC
Total Hydrolyzable Sulfur	3	IC
Carbonate	3	Non-Dispersive Infrared
Trace Metals	1	ICPES
Solid-Phase Analyses		
Wt.% Solids	3	Gravimetric
Acid Insolubles	3	Gravimetric
Calcium	3	Acid Dissolution/AAS
Magnesium	3	Acid Dissolution/AAS
Sulfite	3	Iodometric Titration
Sulfate	3	Acid Dissolution/IC
Carbonate	3	Titration
Morphology	1	Scanning Electron Microscope
TESTS 2 - 7		
Liquid-Phase Analyses		
Calcium	1	AAS
Magnesium	1/Day	AAS
Sodium	1/Day	AAS
Chloride	1/Day	IC
Sulfite	1	Iodometric Titration
Sulfate	1	IC
Total Hydrolyzable Sulfur	1	IC
Carbonate	1	Non-Dispersive Infrared
Solid-Phase Analyses		
Wt.% Solids	3	Gravimetric
Acid Insolubles	1	Gravimetric
Calcium	1	Acid Dissolution/AAS
Magnesium	1	Acid Dissolution/AAS
Sulfite	1	Iodometric Titration
Sulfate	1	Acid Dissolution/IC
Carbonate	3	Titration

Table A-3

Results of Baseline Solids Analyses

Description:	1-U-1*	1-L-1*	1-U-2	1-L-2	1-U-3	1-L-3	2-U-2
Date	9/29/92	9/29/92	9/29/92	9/29/92	9/29/92	9/29/92	9/30/92
Time	1330	1315	1450	1430	1622	1615	1330
Ca, mm/g	5.82	6.54	5.80	5.69	5.86	5.61	6.15
Mg, mm/g	0.01	0.00	0.01	0.00	0.01	0.03	0.02
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.38	5.59	5.36	5.62	5.40	5.53	4.92
SO ₄ , mm/g	5.38	5.27	5.36	5.62	5.40	5.53	4.92
CO ₃ , mm/g	0.43	0.04	0.48	0.05	0.51	0.05	1.42*
Inerts, wt.% of solids	0.40	1.22	0.73	1.02	0.51	1.52	0.80
Solids, wt.% of slurry	18.8	17.3	19.4	17.3	19.7	17.2	19.7*
pH	5.70	3.68	5.73	4.07	5.80	4.07	6.19
Temperature, °C	54.9	51.5	55.0	54.4	55.1	54.7	54.6
Reagent Utilization, % Ca-Independent SO ₄ -Independent	92.7 92.7	99.3 99.4	91.8 91.7	99.0 99.1	91.3 91.3	99.2 99.2	77.6 77.0
Reagent Ratio Ca-Independent SO ₄ -Independent	1.08 1.08	1.01 1.01	1.09 1.09	1.01 1.01	1.09 1.10	1.01 1.01	1.29 1.30
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt.%	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt.%	92.6	96.1	92.2	96.7	92.8	95.0	84.7
CaCO ₃ , wt.%	4.3	0.4	4.8	0.5	5.1	0.5	14.2
Inerts, wt.%	0.4	1.2	0.7	1.0	0.5	1.5	0.8
Ca, mg/g	233	262	232	228	234	224	246
Mg, mg/g	0	0	0	0	0	1	0
SO ₃ , mg/g	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	517	536	515	540	518	530	473
SO ₄ , mg/g	517	506	515	540	518	530	473
CO ₃ , mg/g	26	2	29	3	31	3	85
Closures							
Weight, %	-2.7	1.4	-2.4	-1.7	-1.8	-2.8	-1.1
Molar, %	0.2	7.5	-0.3	0.2	-0.3	0.6	-2.4
Acceptable, %	6.8	7.17	6.8	7.0	6.8	7.0	6.4

Table A-3

(Continued)

Description:	2-L-2	3-U-2	3-L-2	4-U-2	4-L-2	5-U-2	5-L-2
Date	9/30/92	9/30/92	9/30/92	10/01/92	10/01/92	10/01/92	10/01/92
Time	1330	1850	1840	1040	1030	1525	1515
Ca, mm/g	5.60	6.26	6.07	5.72	5.59	6.08	5.77
Mg, mm/g	0.00	0.02	0.03	0.00	0.00	0.01	0.00
SO ₃ , mm/g	0.00	0.00	0.49	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.59	4.95	4.98	5.71	5.78	5.23	5.66
SO ₄ , mm/g	5.29	4.95	4.49	5.71	5.78	5.23	5.66
CO ₃ , mm/g	0.05 ^c	1.31 ^c	1.22 ^c	0.15 ^c	0.01 ^c	0.70 ^c	0.06 ^c
Inerts, wt. % of solids	1.31	0.76	1.52	0.95	1.22	0.12	0.73
Solids, wt. % of slurry	18.1 ^c	19.1 ^c	18.4 ^c	17.8 ^c	18.7 ^c	19.6 ^c	19.6 ^c
pH	3.51	6.18	5.13	4.93	2.92	5.51	3.85
Temperature, °C	54.1	55.4	54.9	55.0	54.5	55.8	53.4
Reagent Utilization, %							
Ca-Independent	99.1	79.1	80.3	97.4	99.9	88.2	98.9
SO ₄ -Independent	99.1	79.1	80.0	97.4	99.9	88.5	98.9
Reagent Ratio							
Ca-Independent	1.01	1.26	1.25	1.03	1.00	1.13	1.01
SO ₄ -Independent	1.01	1.26	1.25	1.03	1.00	1.13	1.01
Oxidation, %	100.0	100.0	90.2	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	7.6	0.0	0.0	0.0	0.0
Gypsum, wt. %	96.1	85.1	75.7	98.2	99.4	90.0	97.4
CaCO ₃ , wt. %	0.5	13.1	12.2	1.5	0.1	7.0	0.6
Inerts, wt. %	1.3	0.8	1.5	1.0	1.2	0.1	0.7
Ca, mg/g	224	250	243	229	224	243	231
Mg, mg/g	0	0	1	0	0	0	0
SO ₃ , mg/g	0	0	39	0	0	0	0
Total S (as SO ₄), mg/g	536	475	478	548	555	502	543
SO ₄ , mg/g	508	475	431	548	555	502	543
CO ₃ , mg/g	3	79	73	9	0	42	4
Closures							
Weight, %	-2.2	-1.0	-3.5	0.1	-0.1	-2.3	-1.1
Molar, %	-0.3	0.2	-0.8	-1.2	-1.7	1.3	0.4
Acceptable, %	7.0	6.4	6.2	7.0	7.1	6.7	7.0

Table A-3

(Continued)

Description:	6-U-2	6-L-2	7-U-2	7-L-2
Date	10/02/92	10/02/92	10/02/92	10/02/92
Time	1130	1120	1600	1550
Ca, mm/g	6.23	6.09	6.06	6.09
Mg, mm/g	0.02	0.02	0.01	0.02
SO ₃ , mm/g	0.00	0.28	0.00	0.64
Total S (as SO ₄), mm/g	4.97	5.16	5.25	5.39
SO ₄ , mm/g	4.97	4.88	5.25	4.75
CO ₃ , mm/g	1.33 ^c	1.09 ^c	0.73 ^c	0.85 ^c
Inerts, wt.% of solids	0.36	0.39	0.32	0.74
Solids, wt.% of slurry	18.4 ^c	19.9 ^c	18.5	20.4 ^c
pH	6.12	6.00	5.91	5.24
Temperature, °C	56.3	55.8	55.8	52.1
Reagent Utilization, % Ca-Independent SO ₄ -Independent	78.9 78.7	82.6 82.1	87.8 88.0	86.4 86.1
Reagent Ratio Ca-Independent SO ₄ -Independent	1.27 1.27	1.21 1.22	1.14 1.14	1.16 1.16
Oxidation, %	100.0	94.6	100.0	88.1
Solid Solution, wt.%	0.0	4.3	0.0	9.9
Gypsum, wt.%	85.4	83.2	90.2	79.8
CaCO ₃ , wt.%	13.3	10.9	7.3	8.5
Inerts, wt.%	0.4	0.4	0.3	0.7
Ca, mg/g	249	244	242	244
Mg, mg/g	0	0	0	0
SO ₃ , mg/g	0	22	0	51
Total S (as SO ₄), mg/g	477	496	504	518
SO ₄ , mg/g	477	469	504	456
CO ₃ , mg/g	80	65	44	51
Closures Weight, % Molar, % Acceptable, %	-1.1 0.4 6.4	-1.8 -1.2 6.4	-1.8 0.8 6.7	-1.6 -1.1 6.3

^a1-U-1 designates Test 1, Upper Loop, Sample 1.

^b1-L-1 designates Test 1, Lower Loop, Sample 1.

^cIndicated wt.% solids and solid carbonate content for Tests 2-7 are an average value for three separate slurry samples.

Table A-4

Results of Baseline Liquid Analyses

Description:	1-U-1	1-U-2	1-U-3	1-L-1	1-L-2	1-L-3	2-U-1	2-L-1	3-U-1	3-L-1
Date	9/29/92	9/29/92	9/29/92	9/29/92	9/29/92	9/29/92	9/30/92	9/30/92	9/30/92	9/30/92
Time	1330	1450	1622	1315	1430	1615	1330	1320	1850	1840
Ca, mm/L	23.4	24.7	24.8	46.1	45.5	46.3	26.1	50.2	27.0	54.8
Mg, mm/L	50.4		58.0	158.4		155.6	57.0	196.7		
Na, mm/L	35.2		35.2	103.9		102.3	34.5	123.8		
Cl, mm/L	115.0	115.0	114.0	423.6	404.0	412.0	132.1	455.0	133.0	449.0
CO ₃ , mm/L	2.34	2.16	2.33	1.95	1.20	1.86	3.93	1.94	4.51	2.08
SO ₃ , mm/L	0.00	0.00	0.00	1.91	1.82	2.06	0.00	1.57	0.00	17.00
SO ₄ , mm/L	39.8	39.6	39.8	45.9	44.6	44.1	38.8	45.7	39.8	48.4
Total Hyd SO ₄ , mm/L	47.7	48.2	46.4	66.1	69.7	70.6	51.3	69.8	48.3	79.2
S/N, mm/L	7.9	8.6	6.6	18.3	23.3	24.4	12.5	22.5	8.5	13.8
pH	5.70	5.73	5.80	3.68	4.07	4.07	6.18	3.51	6.18	5.13
Temperature, °C	54.9	55.0	55.1	51.5	54.4	54.7	54.6	54.1	55.4	54.9
Ca, mg/L	938	989	994	1848	1823	1854	1047	2012	1082	2196
Mg, mg/L	1225		1411	3850		3782	1387	4782		
Na, mg/L	809		809	2388		2352	793	2847		
Cl, mg/L	4077	4077	4041	15017	14322	14605	4683	16130	4715	15917
CO ₃ , mg/L	140	130	140	117	72	112	236	116	271	125
SO ₃ , mg/L	0	0	0	153	146	165	0	126	0	1361
SO ₄ , mg/L	3823	3804	3823	4409	4284	4236	3727	4390	3823	4649
Charge Imbalance										
Calculated, %	-5.2		0.0	-2.2		-2.0	-5.4	4.0		
Acceptable, %	4.8		4.8	5.4		5.4	4.9	5.4		
Relative Saturation										
Gypsum	1.10	1.11	1.09	1.07	1.04	1.04	1.10	1.00	1.16	1.12
CaSO ₃ •0.5H ₂ O	0.00	0.00	0.00	0.07	0.06	0.07	0.00	0.02	0.00	6.00
CaCO ₃	0.04	0.04	0.06	0.00	0.00	0.00	0.36	0.00	0.44	0.01

Table A-4

(Continued)

Description:	4-U-1	4-L-1	5-U-1	5-L-1	6-U-1	6-L-1	7-U-1	7-L-1
Date	10/1/92	10/1/92	10/1/92	10/1/92	10/2/92	10/2/92	10/2/92	10/2/92
Time	1040	1030	1525	1515	1012	1120	1600	1550
Ca, mm/L	26.6	54.4	29.9	56.2	26.7	44.3	27.9	53.3
Mg, mm/L	66.7	183.9			52.7	104.5		
Na, mm/L	40.7	128.0			31.6	64.6		
Cl, mm/L	130.0	503.0	141.0	453.0	107.0	255.0	110.0	392.0
CO ₃ , mm/L	2.17	1.93	2.25	1.89	3.25	3.26	2.54	4.23
SO ₃ , mm/L	0.00	1.43	2.65	2.92	0.55	11.14	0.48	8.15
SO ₄ , mm/L	40.9	44.7	40.2	39.3	34.5	37.5	36.9	42.2
Total Hyd SO ₄ , mm/L	51.8	75.1	52.3	78.0	42.8	67.5	45.5	73.7
S/N, mm/L	10.9	28.9	9.4	35.7	7.8	18.9	8.1	23.4
pH	4.93	2.92	5.51	3.89	6.12	5.34	5.91	5.11
Temperature, °C	55.0	54.5	55.8	53.4	56.3	55.8	55.8	52.1
Ca, mg/L	1067	2181	1200	2251	1072	1776	1116	2136
Mg, mg/L	1621	4471			1280	2541		
Na, mg/L	936	2943			726	1486		
Cl, mg/L	4609	17831	4998	16059	3793	9040	3900	13896
CO ₃ , mg/L	130	116	135	113	195	196	152	254
SO ₃ , mg/L	0	114	212	234	44	892	38	652
SO ₄ , mg/L	3929	4294	3862	3775	3314	3602	3545	4054
Charge Imbalance								
Calculated, %	1.0	-1.5			1.3	0.2		
Acceptable, %	4.9	5.4			4.8	5.1		
Relative Saturation								
Gypsum	1.09	1.06	1.16	1.00	1.07	1.11	1.17	1.10
CaSO ₃ •0.5H ₂ O	0.00	0.00	1.57	0.08	0.74	5.92	0.53	2.94
CaCO ₃	0.00	0.00	0.02	0.00	0.28	0.02	0.11	0.01

Table A-5

**Trace Species
Analytical Data Summary (Baseline Tests)**
(Results are in mg/L.)

Description:	1-U-2		1-U-2 Dup		1-L-2		Method Blank	
Element	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit
Aluminum	1.4	0.6	1.4	0.6	1.8	0.6	ND	0.20
Antimony	0.57	0.30	0.63	0.30	1.5	0.30	ND	0.10
Arsenic	ND	0.9	ND	0.9	ND	0.9	ND	0.30
Barium	0.33	0.03	0.33	0.03	0.70	0.03	ND	0.010
Beryllium	ND	0.006	ND	0.006	ND	0.006	ND	0.0020
Boron	360	7.3	390	7.3	7320	9.0	ND	0.60
Cadmium	0.17	0.0015	0.17	0.0015	0.40	0.0015	ND	0.0050
Calcium	970	3.0	970	3.0	1710	3.0	ND	1.0
Chromium	ND	0.03	ND	0.03	0.11	0.03	ND	0.010
Cobalt	0.054	0.03	0.048	0.03	0.14	0.03	ND	0.010
Copper	ND	0.06	ND	0.06	0.19	0.06	ND	0.020
Iron	ND	0.15	ND	0.15	1.1	0.15	ND	0.050
Lead	ND	0.15	ND	0.15	ND	0.75	ND	0.050
Magnesium	1480	3.0	1480	3.0	3980	15.0	ND	1.0
Manganese	9.4	0.03	9.4	0.03	24.1	0.15	ND	0.010
Molybdenum	ND	0.15	ND	0.15	0.22	0.15	ND	0.050
Nickel	0.36	0.06	0.33	0.06	0.95	0.06	ND	0.020
Potassium	45	9.0	45	9.0	129	9.0	ND	3.0
Selenium	ND	0.90	ND	0.90	1.8	0.90	ND	0.30
Silicon	23.0	3.0	23.0	3.0	61	3.0	ND	1.0
Silver	ND	0.03	ND	0.03	ND	0.03	ND	0.010
Sodium	820	3.0	820	3.0	2330	3.0	ND	1.0
Strontium	4.8	0.009	4.8	0.009	11.6	0.009	ND	0.0030
Thallium	0.85	0.30	0.45	0.30	0.89	0.30	ND	0.10
Vanadium	ND	0.06	ND	0.06	ND	0.30	ND	0.020
Zinc	ND	0.06	ND	0.06	ND	0.30	ND	0.020

ND = not detected at the specified detection limit.

APPENDIX B

Results of Chemical Analyses for the Parametric and Additive Consumption Tests

Table B-1

Actual Parametric Test Conditions and Results

Test No.	Run No.	Date (1992)	Upper Loop pH *	Lower Loop pH *	Upper Loop DBA (ppm)	Lower Loop DBA (ppm)	Flue Gas Velocity (ft/s)	Outlet Flue Gas O ₂ (%)	Lear Siegler Inlet SO ₂ (ppmv)	M6 Lower Loop SO ₂ (ppmv)	M6 Outlet SO ₂ (ppmv)	DuPont Outlet SO ₂ (ppmv)	SO ₂ Removal Efficiencies (%)	
													Lower Loop	Overall
1	1	11/04	5.9/5.9	3.5/3.5	93	443	7.6	5.6	2597	2953	155	139	c	94.0
1	2	11/04	5.9/5.9	3.5/3.5	93	443	7.6	5.6	2553	2717	156	133	c	93.9
1	3	11/04	5.9/5.9	3.8/3.8	47	169	7.6	5.6	2531	2694	143	112	c	94.4
1	4	11/04	5.9/5.9	3.8/3.8	47	169	7.6	5.6	2540	2736	158	128	c	93.8
1	5	11/04	5.8/5.8	3.4/3.3	70	439	7.6	5.6	2535	2482	159	141	2	93.7
1	6	11/04	5.8/5.8	3.4/3.3	70	439	7.6	5.6	2487	2466	170	124	1	93.1
1	7	11/05	5.8/5.8	3.6/3.5	45	121	8.1	5.9	2480	2536	203	168	c	91.8
1	8	11/05	5.9/5.9	3.5/3.5	114	432	8.1	5.9	2520	2394	146	129	5	94.2
2	1	11/05	5.9/5.8	3.6/3.5	88	338	10.1	6.8	2556	2724	272	175	c	89.1
2	2	11/05	5.9/5.8	3.6/3.5	88	338	10.1	6.8	2558	2790	302	188	c	88.2
2	3	11/05	5.8/5.8	3.7/3.7	69	149	10.1	6.8	2560	2680	304	NA	c	88.1
2	4	11/05	5.9/5.8	3.6/3.5	69	149	10.1	6.8	2580	2727	322	NA	c	87.5
3	1	11/06	6.3/6.2	3.6/3.5	48	351	7.8	6	2795	2401	85	87	14	97.0
3	2	11/06	6.2/6.1	3.8/3.8	48	351	7.8	6	2793	2489	90	92	11	96.8
4	1	11/06	6.0/6.0	3.7/3.5	83	319	10.1	6.2	2848	2408	199	145	16	93.0
4	2	11/06	6.0/5.9	3.7/3.6	83	319	10.1	6.2	2807	2756	203	147	2	92.8
5	1	11/08	5.9/5.9	3.5/3.6	214	1155	7.4	7	2630	2672	105	118	c	96.0
5	2	11/08	5.9/5.9	3.5/3.6	214	1155	7.4	7	2528		88	141	b	96.5
5	3	11/08	5.8/5.8	3.7/3.6	215	915	7.4	6.5	2552		69	145	b	97.3
5	4	11/08	5.8/5.8	3.8/3.9	215	915	7.4	6.5	2546		71	119	b	97.2
5	5	11/12	5.8/5.8	3.5/3.5	230	509	7.4	7.0	2720	2457	71	62	10	97.4
5	6	11/12	5.8/5.8	3.5/3.5	230	509	8.15	7.0	2640	2372	63	57	10	97.6

Table B-1

(Continued)

Test No.	Run No.	Date (1992)	Upper Loop pH *	Lower Loop pH *	Upper Loop DBA (ppm)	Lower Loop DBA (ppm)	Flue Gas Velocity (ft/s)	Outlet Flue Gas O ₂ (%)	Lear Siegler Inlet SO ₂ (ppmv)	M6 Lower Loop SO ₂ (ppmv)	M6 Outlet SO ₂ (ppmv)	DuPont Outlet SO ₂ (ppmv)	SO ₂ Removal Efficiencies (%)	
													Lower Loop	Overall
5	7	11/12	5.8/5.8	3.4/3.4	157	366	7.6	5.9	2619	2444	87	74	7	96.7
5	8	11/12	5.9/5.8	3.4/3.4	157	366	7.6	5.9	2633	2483	89	78	6	96.6
6	1	11/13	5.9/5.8	3.4/3.4	176	402	9.5	6.2	2716	2296	165	129	16	93.9
6	2	11/13	5.9/5.8	3.4/3.4	176	402	9.5	6.2	2721	2507	157	129	8	94.2
6	3	11/13	5.8/5.7	3.4/3.4	140	407	9.5	7.5	2779	2423	172	139	13	93.8
6	4	11/13	5.8/5.8	3.5/3.5	140	407	9.5	7.5	2772	2582	158	131	7	94.3
7	1	11/14	6.1/6.1	3.2/3.2	117	532	7.3	6	2623	2411	62	58	8	97.6
7	2	11/14	6.1/6.1	3.7/3.6	117	532	7.3	6	2604	2499	65	65	4	97.5
8	1	11/14	6.1/6.1	3.6/3.7	120	347	9.7	7	2828	2904	170	116	c	94.0
8	2	11/14	6.1/6.1	4.1/4.3	120	347	9.7	7	2793	2765	191	120	1	93.2
9	1	11/15	5.9/5.9	3.5/3.5	277	1208	7.5	6.8	2597	2387	65	75	8	97.5
9	2	11/15	5.9/5.9	3.5/3.5	277	1208	7.5	6.8	2681	2517	71	63	6	97.3
9	3	11/15	5.9/5.9	3.5/3.6	213	1535	7.5	6.8	2754	2671	71	69	3	97.4
9	4	11/15	5.8/5.8	3.5/3.5	213	1535	7.5	6.8	2759	2480	87	78	10	96.8
9	5	11/16	5.9/5.9	3.5/3.4	365	843	7.2	10.2	2636	2188	25	54	17	99.0
9	6	11/16	5.9/5.9	3.5/3.4	365	843	7.2	10.2	2675	2216	29	35	17	98.9
9	7	11/16	5.8/5.8	3.4/3.5	309	986	7.2	6.5	2752	3030	54	44	c	98.0
9	8	11/16	5.8/5.8	3.4/3.5	309	986	7.2	6.5	2702	2963	46	45	c	98.3
10	1	11/17	5.9/5.8	3.5/3.6	417	1217	9.8	8.2	2660	2593	62	52	3	97.7
10	2	11/17	5.9/5.8	3.5/3.6	417	1217	9.8	8.2	2670	2651	62	58	1	97.7
11	1	11/17	5.7/5.8	3.3/3.6	441	1023	9.2	7.9	2634	2346	30	44	11	98.9

Table B-1

(Continued)

Test No.	Run No.	Date (1992)	Upper Loop pH ^a	Lower Loop pH ^a	Upper Loop DBA (ppm)	Lower Loop DBA (ppm)	Flue Gas Velocity (ft/s)	Outlet Flue Gas O ₂ (%)	Lear Siegler Inlet SO ₂ (ppmv)	M6 Lower Loop SO ₂ (ppmv)	M6 Outlet SO ₂ (ppmv)	DuPont Outlet SO ₂ (ppmv)	SO ₂ Removal Efficiencies (%)	
													Lower Loop	Overall
11	2	11/17	5.8/5.8	3.4/3.6	441	1023	9.2	7.9	2640	2353	26	30	11	99.0
12	1	11/18	6.1/6.1	3.7/3.7	470	1172	7.5	9	2768	2281	16	na	18	99.4
12	2	11/18	6.1/6.1	3.7/3.7	470	1172	7.5	9	2702	2248	18	70	17	99.3
13	1	11/18	6.1/6.0	3.7/3.7	423	1180	9.8	7.7	2756	2671	49	38	3	98.2
13	2	11/18	6.0/6.0	3.7/3.7	423	1180	9.8	7.7	2737	2290	51	21	16	98.1
14	1	11/19	6.1/6.1	3.6/3.6	907	2137	7.75	8.3	2927	1895	9	17	35	99.7
14	2	11/19	6.0/6.0	3.6/3.6	907	2137	7.75	8.3	2890	2217	9	20	23	99.7
15	1	11/19	6.1/6.0	3.6/3.6	881	1800	9.9	8	2901	2669	26	37	8	99.1
15	2	11/19	5.9/6.0	3.5/3.6	881	1800	9.9	8	2883	2721	25	48	18.6	99.1

^aReported pH values are those measured by Radian and by the local on-line pH indicator, respectively.

^bMidpoint samples lost due to probe leak.

^cMidpoint SO₂ concentration greater than inlet concentration.

Table B-2

Chemical Analyses Conducted for Parametric Tests

	Number of Analyses (Each Loop)	Technique
TEST 1, 5, and 9		
Liquid-Phase Analyses		
Calcium	4	Atomic Absorption Spectroscopy (AAS)
Magnesium	2	AAS
Sodium	2	AAS
Chloride	2	Ion Chromatography (IC)
Sulfite	4	Iodometric Titration
Sulfate	4	IC
Total Hydrolyzable Sulfur	4	IC
Carbonate	4	Non-Dispersive Infrared
Trace Metals	1	ICPES
DBA	4	ICE
Solid-Phase Analyses		
Wt.% Solids	6	Gravimetric
Acid Insolubles	2	Gravimetric
Calcium	2	Acid Dissolution/AAS
Magnesium	2	Acid Dissolution/AAS
Sulfite	2	Iodometric Titration
Sulfate	2	Acid Dissolution/IC
Carbonate	6	Titration
Morphology	1	Scanning Electron Microscope

Table B-2

(Continued)

	Number of Analyses (Each Loop)	Technique
TESTS 2 and 6		
Liquid-Phase Analyses		
Calcium	2	AAS
Magnesium	1	AAS
Sodium	1	AAS
Chloride	1	IC
Sulfite	2	Iodometric Titration
Sulfate	2	IC
Total Hydrolyzable Sulfur	2	IC
Carbonate	2	Non-Dispersive Infrared
Trace Metals	1	ICPES
DBA	2	ICE
Solid-Phase Analyses		
Wt.% Solids	3	Gravimetric
Acid Insolubles	2	Gravimetric
Calcium	2	Acid Dissolution/AAS
Magnesium	2	Acid Dissolution/AAS
Sulfite	2	Iodometric Titration
Sulfate	2	Acid Dissolution/IC
Carbonate	3	Titration
TESTS 3, 4, 7, 8, 10, 11, 12, 13, 14, and 15		
Liquid-Phase Analyses		
Calcium	1	AAS
Magnesium	1/Day	AAS
Sodium	1/Day	AAS
Chloride	1/Day	IC
Sulfite	1	Iodometric Titration
Sulfate	1	IC
Total Hydrolyzable Sulfur	1	IC
Carbonate	1	Non-Dispersive Infrared
DBA	1	ICE
Solid-Phase Analyses		
Wt.% Solids	3	Gravimetric
Acid Insolubles	1	Gravimetric
Calcium	1	Acid Dissolution/AAS
Magnesium	1	Acid Dissolution/AAS
Sulfite	1	Iodometric Titration
Sulfate	1	Acid Dissolution/IC
Carbonate	3	Titration

Table B-2

(Continued)

	Number of Analyses (Each Loop)	Technique
LONG-TERM TESTING		
Liquid-Phase Analyses		
Calcium	3	AAS
Magnesium	3	AAS
Sodium	3	AAS
Chloride	3	IC
Sulfite	3	Iodometric Titration
Sulfate	3	IC
Total Hydrolyzable Sulfur	3	IC
Carbonate	3	Non-Dispersive Infrared
Trace Metals	1	ICPES
DBA	See Table 5 *	ICE
Solid-Phase Analyses		
Wt.% Solids	3	Gravimetric
Acid Insolubles	3	Gravimetric
Calcium	3	Acid Dissolution/AAS
Magnesium	3	Acid Dissolution/AAS
Sulfite	3	Iodometric Titration
Sulfate	3	Acid Dissolution/IC
Carbonate	3	Titration
Morphology	1	Scanning Electron Microscope

* DBA samples were obtained from all vessels in the FGD system.

Table B-3

Results of Parametric Solids Analyses

Description:	1-U-1	1-L-1	1-U-6	1-L-6	2-U-1	2-L-1	2-U-3	2-L-3
Date	11/04/92	11/04/92	11/05/92	11/05/92	11/05/92	11/05/92	11/05/92	11/05/92
Time	0930	0915	0900	0840	1505	1445	1630	1605
Ca, mm/g	5.73	5.57	5.83	5.64	5.86	5.57	5.77	5.50
Mg, mm/g	0.01	0.00	0.01	0.00	0.01	0.00	0.01	0.00
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.88	6.08	5.85	6.46	5.80	6.11	5.72	6.21
SO ₄ , mm/g	5.88	6.08	5.85	6.46	5.80	6.11	5.72	6.21
CO ₃ , mm/g	0.44	0.02	0.49	0.01	0.48	0.02	0.42	0.01
Inerts, wt. %								
Solids, wt. %	21.4	18.1	22.0	17.9	21.0	17.8	20.9	17.8
pH	5.85	3.59	5.85	3.60	5.87	3.68	5.88	3.67
Temperature, °C	56.8	56.0	56.8	54.8	57.1	56.3	56.9	55.8
Reagent Utilization, % Ca-Independent SO ₄ -Independent	93.0 92.3	99.7 99.7	92.3 91.6	99.8 99.8	92.4 91.8	99.6 99.6	93.2 92.7	99.8 99.8
Reagent Ratio Ca-Independent SO ₄ -Independent	1.07 1.08	1.00 1.00	1.08 1.09	1.00 1.00	1.08 1.09	1.00 1.00	1.07 1.08	1.00 1.00
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt. %	101.1	104.6	100.6	111.1	99.8	105.1	98.4	106.8
CaCO ₃ , wt. %	4.4	0.2	4.9	0.1	4.8	0.2	4.2	0.1
Inerts, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ca, mg/g	229	223	233	226	234	223	231	220
Mg, mg/g	0	0	0	0	0	0	0	0
SO ₃ , mg/g	0	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	564	584	562	620	557	586	549	596
SO ₄ , mg/g	564	584	562	620	557	586	549	596
CO ₃ , mg/g	26	1	29	1	29	1	25	1
Closures								
Weight, %	3.2	2.7	3.5	7.9	2.9	3.1	1.2	4.1
Molar, %	-4.8	-4.5	-4.1	-6.9	-3.4	-4.8	-3.0	-6.2
Acceptable, %	6.8	7.1	6.8	7.1	6.8	7.1	6.8	7.1

Table B-3

(Continued)

Description:	3-U-2	3-L-2	4-U-2	4-L-2	5-U-1	5-L-1	5-U-3	5-L-3
Date	11/6/92	11/6/92	11/6/92	11/6/92	11/8/92	11/8/92	11/8/92	11/8/92
Time	1150	1130	1830	1815	1100	1030	1520	1500
Ca, mm/g	5.93	5.57	5.97	5.48	5.68	5.63	5.74	5.56
Mg, mm/g	0.02	0.00	0.02	0.01	0.01	0.00	0.01	0.00
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.60	6.07	5.63	6.00	5.88	6.18	5.88	6.13
SO ₄ , mm/g	5.60	6.07	5.63	6.00	5.88	6.18	5.88	6.13
CO ₃ , mm/g	1.35	0.02	0.92	0.02	0.39	0.01	0.39	0.02
Inerts, wt. %								
Solids, wt. %	21.7	18.3	24.4	17.6	22.9	20.3	22.0	16.8
pH	6.18	3.78	5.97	3.59	5.87	3.60	5.83	3.74
Temperature, °C	56.0	55.4	56.5	55.7	55.6	55.8	56.2	55.6
Reagent Utilization, % Ca-Independent SO ₄ -Independent	80.6 77.3	99.7 99.6	86.0 84.6	99.7 99.6	93.8 93.1	99.9 99.9	93.8 93.2	99.8 99.7
Reagent Ratio Ca-Independent SO ₄ -Independent	1.24 1.29	1.00 1.00	1.16 1.18	1.00 1.00	1.07 1.07	1.00 1.00	1.07 1.07	1.00 1.00
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt. %	96.3	104.3	96.8	103.2	101.1	106.4	101.1	105.4
CaCO ₃ , wt. %	13.5	0.2	9.2	0.2	3.9	0.1	3.9	0.2
Inerts, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ca, mg/g	237	223	239	219	227	225	230	222
Mg, mg/g	1	0	0	0	0	0	0	0
SO ₃ , mg/g	0	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	538	582	540	576	564	594	564	588
SO ₄ , mg/g	538	582	540	576	564	594	564	588
CO ₃ , mg/g	81	1	55	1	23	0	23	1
Closures								
Weight, %	5.8	2.5	3.8	1.3	2.7	4.2	2.9	3.2
Molar, %	-7.7	-4.4	-4.5	-4.7	-4.9	-4.7	-4.3	-4.9
Acceptable, %	6.4	7.1	6.6	7.1	6.8	7.1	6.8	7.1

Table B-3

(Continued)

Description:	5-U-6	5-L-6	6-U-1	6-L-1	6-U-3	6-L-3	7-U-2	7-L-2
Date	11/12/92	11/12/92	11/13/92	11/13/92	11/13/92	11/13/92	11/14/92	11/14/92
Time	1400	1420	0930	0930	1400	1340	1000	0945
Ca, mm/g	5.44	5.61	5.40	5.19	5.67	5.09	6.04	5.69
Mg, mm/g	0.01	0.00	0.00	0.00	0.00	0.01	0.00	0.05
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.47	6.10	5.53	5.80	5.78	5.53	5.53	6.18
SO ₄ , mm/g	5.47	6.10	5.53	5.80	5.78	5.53	5.53	6.18
CO ₃ , mm/g	0.49	0.02	0.50	0.02	0.49	0.02	1.06	0.02
Inerts, wt. %								
Solids, wt. %	22.8	16.6	23.1	20.2	20.4	18.1	15.9	16.4
pH	5.79	3.50	5.87	3.44	5.85	3.5	6.12	3.51
Temperature, °C	56.7	57.3	56.8	56.3	57.5	57	55.6	55.6
Reagent Utilization, % Ca-Independent SO ₄ -Independent	91.7 91.0	99.7 99.6	91.7 90.7	99.6 99.6	92.2 91.4	99.7 99.7	83.9 82.5	99.7 99.7
Reagent Ratio Ca-Independent SO ₄ -Independent	1.09 1.10	1.00 1.00	1.09 1.10	1.00 1.00	1.08 1.09	1.00 1.00	1.19 1.21	1.00 1.00
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt. %	94.0	104.9	95.2	99.8	99.4	95.1	95.1	106.2
CaCO ₃ , wt. %	4.9	0.2	5.0	0.2	4.9	0.2	10.6	0.2
Inerts, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ca, mg/g	218	224	216	208	227	204	242	228
Mg, mg/g	0	0	0	0	0	0	0	1
SO ₃ , mg/g	0	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	525	586	531	557	555	531	531	593
SO ₄ , mg/g	525	586	531	557	555	531	531	593
CO ₃ , mg/g	30	1	30	1	29	1	64	1
Closures								
Weight, %	-3.1	3.1	-2.4	-2.5	1.9	-6.5	3.5	4.5
Molar, %	-4.5	-4.3	-5.5	-5.7	-5.0	-4.2	-4.3	-3.9
Acceptable, %	6.8	7.1	6.8	7.1	6.8	7.1	6.5	7.0

Table B-3

(Continued)

Description:	8-U-2	8-L-2	9-U-1	9-L-1	9-U-6	9-L-6	10-U-2	10-L-2
Date	11/14/92	11/14/92	11/15/92	11/15/92	11/16/92	11/16/92	11/17/92	11/17/92
Time	1430	1410	1045	1030	1520	1500	1020	1000
Ca, mm/g	6.20	5.53	5.87	5.58	5.69	5.60	5.73	5.60
Mg, mm/g	0.00	0.00	0.01	0.00	0.01	0.00	0.01	0.01
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.24	6.16	5.78	6.15	5.24	5.45	5.62	6.38
SO ₄ , mm/g	5.24	6.16	5.78	6.15	5.24	5.45	5.62	6.38
CO ₃ , mm/g	1.46	0.05	0.61	0.01	0.61	0.01	0.43	0.01
Inerts, wt. %								
Solids, wt. %	15.4	16.8	13.1	13.6	22.1	19.6	24.5	22.3
pH	6.08	3.9	5.89	3.48	5.87	3.47	5.85	3.49
Temperature, °C	56.1	55.4	55.5	54.8	56.3	55.2	55	54.4
Reagent Utilization, %								
Ca-Independent	78.2	99.2	90.5	99.9	89.6	99.8	92.9	99.9
SO ₄ -Independent	76.5	99.1	89.6	99.9	89.3	99.9	92.5	99.8
Reagent Ratio								
Ca-Independent	1.28	1.01	1.11	1.00	1.12	1.00	1.08	1.00
SO ₄ -Independent	1.31	1.01	1.12	1.00	1.12	1.00	1.08	1.00
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt. %	90.0	106.0	99.5	105.7	90.1	93.7	96.6	109.7
CaCO ₃ , wt. %	14.6	0.5	6.1	0.1	6.1	0.1	4.3	0.1
Inerts, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ca, mg/g	248	221	235	223	228	224	229	224
Mg, mg/g	0	0	0	0	0	0	0	0
SO ₃ , mg/g	0	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	503	592	555	590	503	523	539	612
SO ₄ , mg/g	503	592	555	590	503	523	539	612
CO ₃ , mg/g	88	3	37	0	37	1	26	1
Closures								
Weight, %	2.7	3.7	3.5	3.5	-4.4	-5.7	-0.3	6.7
Molar, %	-3.8	-5.8	-4.2	-4.9	-1.3	1.3	-2.6	-6.5
Acceptable, %	6.4	7.1	6.7	7.1	6.7	7.1	6.8	7.1

Table B-3

(Continued)

Description:	11-U-2	11-L-2	12-U-2	12-L-2	13-U-2	13-L-2	14-U-2	14-L-2
Date	11/17/92	11/17/92	11/18/92	11/18/92	11/18/92	11/18/92	11/19/92	11/19/92
Time	1430	1415	1110	1050	1600	1600	1100	1045
Ca, mm/g	5.80	5.50	5.79	5.54			5.94	5.66
Mg, mm/g	0.01	0.02	0.01	0.00			0.01	0.00
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.84	6.15	5.78	6.15			5.39	5.89
SO ₄ , mm/g	5.84	6.15	5.78	6.15			5.39	5.89
CO ₃ , mm/g	0.36	0.02	0.83	0.01	0.63	0.01	0.82	0.01
Inerts, wt. %								
Solids, wt. %	24.8	19.4	19.6	18.7	20.8	18.3	19.9	18.1
pH	5.78	3.67	6.06	3.68	6.04	3.72	6.07	3.63
Temperature, °C	55.5	54.8	53.8	53.4	56.3	54.7	54	53.5
Reagent Utilization, % Ca-Independent SO ₄ -Independent	94.2 93.8	99.7 99.7	87.4 85.7	99.8 99.7			86.9 86.9	99.9 99.9
Reagent Ratio Ca-Independent SO ₄ -Independent	1.06 1.07	1.00 1.00	1.14 1.17	1.00 1.00			1.15 1.16	1.00 1.00
Oxidation, %	100.0	100.0	100.0	100.0			100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0			0.0	0.0
Gypsum, wt. %	100.5	105.7	99.3	105.8			92.7	101.3
CaCO ₃ , wt. %	3.6	0.2	8.3	0.1	6.3	0.1	8.2	0.1
Inerts, wt. %	0.0	0.0	0.0	0.0			0.0	0.0
Ca, mg/g	232	220	232	222			238	226
Mg, mg/g	0	1	0	0			0	0
SO ₃ , mg/g	0	0	0	0			0	0
Total S (as SO ₄), mg/g	561	590	554	591			517	565
SO ₄ , mg/g	561	590	554	591			517	565
CO ₃ , mg/g	22	1	50	1	38	1	49	1
Closures								
Weight, %	2.5	3.3	4.4	3.5			2.2	1.3
Molar, %	-3.3	-5.5	-6.5	-5.3			-1.2	-2.0
Acceptable, %	6.9	7.1	6.6	7.1			6.6	7.1

Table B-3

(Continued)

Description:	15-U-2	15-L-2	LT-1-UB	LT-1-LB	LT-4-UB	LT-4-LB	LT-5-UB	LT-5-LB
Date	11/19/92	11/19/92						
Time	1645	1630						
Ca, mm/g	6.16	5.70	5.59	5.51	5.63	5.58	5.54	5.53
Mg, mm/g	0.02	0.01	0.01	0.02	0.00	0.00	0.01	0.02
SO ₃ , mm/g	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total S (as SO ₄), mm/g	5.49	5.94	5.57	5.49	5.54	5.59	5.64	5.71
SO ₄ , mm/g	5.49	5.94	5.57	5.49	5.54	5.59	5.64	5.71
CO ₃ , mm/g	0.83	0.01	0.18	0.01	0.19	0.01	0.12	0.01
Inerts, wt. %								
Solids, wt. %	21.2	17.9	19.2	15.3	19.7	15.9	17.3	17.0
pH	6.03	3.61						
Temperature, °C	56.1	58.7						
Reagent Utilization, % Ca-Independent SO ₄ -Independent	86.9 86.6	99.9 99.9	96.9 96.8	99.9 99.9	96.7 96.6	99.9 99.9	97.9 97.8	99.9 99.9
Reagent Ratio Ca-Independent SO ₄ -Independent	1.15 1.16	1.00 1.00	1.03 1.03	1.00 1.00	1.03 1.03	1.00 1.00	1.02 1.02	1.00 1.00
Oxidation, %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Solid Solution, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Gypsum, wt. %	94.5	102.2	95.9	94.5	95.3	96.1	97.1	98.2
CaCO ₃ , wt. %	8.3	0.1	1.8	0.1	1.9	0.1	1.2	0.1
Inerts, wt. %	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ca, mg/g	246	228	223	220	225	223	222	221
Mg, mg/g	0	0	0	0	0	0	0	0
SO ₃ , mg/g	0	0	0	0	0	0	0	0
Total S (as SO ₄), mg/g	527	570	535	527	532	536	542	548
SO ₄ , mg/g	527	570	535	527	532	536	542	548
CO ₃ , mg/g	50	0	11	0	11	0	7	0
Closures								
Weight, %	2.2	1.3	-3.0	-5.4	-3.2	-3.9	-2.6	-2.4
Molar, %	-1.2	-2.0	-1.4	0.2	-0.8	-0.1	-1.9	-1.5
Acceptable, %	6.6	7.1	7.0	7.1	7.0	7.1	7.0	7.1

Table B-4

Results of Parametric Liquid Analyses

Description:	1-U-1	1-L-1	1-U-3	1-L-3	1-U-4	1-L-4	1-U-6	1-L-6	2-U-1	2-L-1
Date	11/4/92	11/4/92	11/4/92	11/4/92	11/4/92	11/4/92	11/5/92	11/5/92	11/5/92	11/5/92
Time	0930	0915	1245	1230	1720	1700	0900	0840	1505	1445
Ca, mm/L	23.4	46.2	23.1	47.3	23.4	46.9	23.3	45.3	23.4	47.1
Mg, mm/L	60.3	133.2					63.4	149.4		
Na, mm/L	10.6	52.5					18.1	50.1		
Cl, mm/L	89.0	327.0					95.2	341.7		
CO ₃ , mm/L	2.95	1.95	2.62	2.09	2.84	1.96	3.08	2.08	2.69	2.22
SO ₃ , mm/L	0.94	1.27	0.29	1.89	1.02	0.39	1.40	0.74	1.92	1.60
SO ₄ , mm/L	43.9	41.1	41.1	42.3	44.9	42.8	47.3	45.0	46.0	45.0
Total Hyd SO ₄ , mm/L	43.7	53.9	47.3	63.1	50.1	57.6	53.8	59.4	54.2	58.7
S/N, mm/L	-1.1	11.5	5.9	19.0	4.1	14.4	5.1	13.7	6.3	12.1
pH	5.88	3.55	5.85	3.85	5.82	3.37	5.85	3.60	5.89	3.60
Temperature, °C	56.8	56.0	56.8	56.0	56.8	56.0	56.8	54.8	57.1	56.3
Ca, mg/L	939	1853	925	1894	938	1879	934	1814	940	1886
Mg, mg/L	1465	3239					1541	3631		
Na, mg/L	243	1206					416	1152		
Cl, mg/L	3154	11592					3376	12112		
CO ₃ , mg/L	177	117	157	125	170	117	185	125	162	133
SO ₃ , mg/L	75	101	23	152	82	31	112	59	153	128
SO ₄ , mg/L	4213	3947	3944	4059	4316	4112	4539	4320	4415	4322
Charge Imbalance										
Calculated, %	0.2	-1.3					-1.5	-0.7		
Acceptable, %	5.1	5.4					5.0	5.4		
Relative Saturation										
Gypsum	1.17	1.13	1.11	1.18	1.19	1.19	1.18	1.14	1.17	1.19
CaSO ₄ •0.5H ₂ O	0.55	0.01	0.17	0.03	0.55	0.00	0.76	0.01	1.11	0.01
CaCO ₃	0.02	0.00	0.01	0.00	0.01	0.00	0.01	0.00	0.01	0.00

Table B-4

(Continued)

Description:	2-U-3	2-L-3	3-U-2	3-L-2	4-U-2	4-L-2	5-U-1	5-L-1	5-U-3	5-L-3
Date	11/5/92	11/5/92	11/6/92	11/6/92	11/6/92	11/6/92	11/8/92	11/8/92	11/8/92	11/8/92
Time	1630	1630	1150	1130	1830	1815	1100	1030	1520	1500
Ca, mm/L	23.2	40.7	23.1	41.9	22.9	47.9	23.3	46.6	23.7	44.1
Mg, mm/L	60.9	143.5	56.0	146.9			64.0	194.1		
Na, mm/L	21.2	53.3	18.2	53.3			15.3	64.2		
Cl, mm/L	87.8	309.5	78.4	337.6			96.9	450.2		
CO ₃ , mm/L	2.90	2.05	3.62	2.24	3.66	1.78	2.90	1.90	3.01	1.96
SO ₃ , mm/L	1.33	1.60	1.07	1.19	2.60	1.71	0.82	1.65	0.69	2.00
SO ₄ , mm/L	43.6	46.0	44.5	46.7	46.2	49.9	46.4	49.9	47.8	51.7
Total Hyd SO ₄ , mm/L	49.1	62.1	53.4	63.0	55.1	69.2	54.4	70.6	57.2	71.6
S/N, mm/L	4.2	14.5	7.9	15.1	6.3	17.6	7.2	19.0	8.7	17.9
pH	5.91	3.59	6.19	3.45	5.95	3.68	5.94	3.54	5.85	3.81
Temperature, °C	56.9	55.8	56.0	55.4	56.5	55.7	55.6	55.8	56.2	55.6
Ca, mg/L	930	1633	927	1679	916	1919	934	1868	952	1769
Mg, mg/L	1479	3489	1362	3571			1555	4718		
Na, mg/L	486	1225	418	1226			351	1475		
Cl, mg/L	3112	10973	2781	11967			3436	15958		
CO ₃ , mg/L	174	123	217	134	219	107	174	114	181	118
SO ₃ , mg/L	107	128	86	95	208	137	65	132	55	160
SO ₄ , mg/L	4189	4418	4272	4486	4442	4793	4457	4793	4593	4970
Charge Imbalance										
Calculated, %	2.1	0.5	-0.3	-1.9			-2.3	-2.2		
Acceptable, %	4.9	5.4	4.9	5.4			5.0	5.6		
Relative Saturation										
Gypsum	1.13	1.09	1.20	1.11	1.21	1.32	1.18	1.09	1.22	1.07
CaSO ₃ •0.5H ₂ O	0.79	0.01	0.88	0.01	1.63	0.02	0.50	0.01	0.38	0.02
CaCO ₃	0.02	0.00	0.06	0.00	0.02	0.00	0.02	0.00	0.01	0.00

Table B-4

(Continued)

Descriptions:	5-U-4	5-L-4	5-U-6	5-L-6	6-U-1	6-L-1	6-U-3	6-L-3	7-U-2	7-L-2
Date	11/12/92	11/12/92	11/12/92	11/12/92	11/13/92	11/13/92	11/13/92	11/13/92	11/14/92	11/14/92
Time	0920	0900	1420	1400	0930	0900	1400	1340	1000	0945
Ca, mm/L	23.4	40.3	23.1	36.4	23.1	37.2	23.1	34.8	21.1	27.3
Mg, mm/L			66.4	140.5			63.8	176.4	33.4	94.8
Na, mm/L			19.1	53.5			12.9	57.8	11.5	30.1
Cl, mm/L			92.4	364.5			88.0	357.3	56.2	164.9
CO ₃ , mm/L			2.40	4.11	3.37	2.92	3.10	2.63	3.56	2.49
SO ₃ , mm/L	0.91	0.47	1.06	0.81	0.82	1.12	0.69	0.78	0.71	1.26
SO ₄ , mm/L	47.7	52.2	44.3	58.3	47.0	59.4	44.0	58.8	36.5	51.4
Total Hyd SO ₄ , mm/L	56.6	73.7	55.5	88.1	55.5	89.2	47.4	87.6	46.1	61.4
S/N, mm/L	8.0	21.1	10.1	29.0	7.7	28.7	2.8	28.0	8.8	8.7
pH	5.75	3.54	5.87	3.39	5.89	3.42	5.85	3.53	6.10	3.48
Temperature, °C	55.9	55.0	57.3	56.7	56.8	56.3	57.5	57.0	55.6	55.6
Ca, mg/L	938	1614	925	1458	928	1492	927	1397	847	1093
Mg, mg/L			1613	3416			1550	4287	812	2304
Na, mg/L			439	1229			297	1328	264	691
Cl, mg/L			3274	12923			3119	12666	1992	5845
CO ₃ , mg/L			144	247	202	175	186	158	214	149
SO ₃ , mg/L	73	38	85	65	66	89	55	62	57	101
SO ₄ , mg/L	4584	5016	4259	5600	4512	5708	4223	5649	3511	4937
Charge Imbalance Calculated, %			1.8	-11.2				-2.3	-6.8	-0.4
Acceptable, %			5.0	5.4			2.2	5.5	4.9	5.1
Relative Saturation Gypsum CaSO ₃ ·0.5H ₂ O	1.18	1.20	1.10	1.20	1.19	1.12	1.14	1.04	1.22	1.14
CaCO ₃	0.42	0.00	0.59	0.00	0.47	0.00	0.38	0.00	0.61	0.00
	0.01	0.00	0.01	0.00	0.02	0.00	0.01	0.00	0.04	0.00

Table B-4

(Continued)

Description:	8-U-2	8-L-2	9-U-1	9-L-1	9-U-3	9-L-3	9-U-4	9-L-4	9-U-6	9-L-6
Date	11/14/92	11/14/92	11/15/92	11/15/92	11/15/92	11/15/92	11/16/92	11/16/92	11/16/92	11/16/92
Time	1430	1410	1045	1030	1300	1245	1045	1030	1520	1500
Ca, mm/L	20.4	29.0	19.5	21.6	19.4	26.7	23.2	32.9	21.4	32.4
Mg, mm/L			13.5	31.3					35.8	122.8
Na, mm/L			7.2	13.5					12.4	43.2
Cl, mm/L			14.4	56.7					40.3	251.5
CO ₃ , mm/L	3.84	2.84	3.11	1.80	3.16	2.31	3.26	2.30	3.52	1.97
SO ₃ , mm/L	1.67	4.34	1.55	0.68	1.97	1.40	1.83	1.34	2.75	1.70
SO ₄ , mm/L	30.2	44.5	24.4	27.4	26.0	31.1	34.6	51.4	34.8	47.0
Total Hyd SO ₄ , mm/L	38.3	57.0	32.0	34.6	33.8	45.1	35.7	71.2	44.3	75.4
S/N, mm/L	6.4	8.1	6.0	6.5	5.8	12.6	-0.8	18.4	6.7	26.8
pH	6.09	4.12	5.95	3.50	5.82	3.45	5.95	3.54	5.8	3.39
Temperature, °C	56.1	55.4	55.5	54.8	55.7	54.9	52.9	51.9	56.3	55.2
Ca, mg/L	819	1163	783	867	779	1070	9292	1320	859	1297
Mg, mg/L			327	761					869	2984
Na, mg/L			165	311					285	992
Cl, mg/L			511	2009					1428	8915
CO ₃ , mg/L	230	170	187	108	190	139	196	138	211	118
SO ₃ , mg/L	133	347	124	54	158	112	146	108	220	136
SO ₄ , mg/L	2897	4273	2345	2630	2499	2987	3325	4940	3340	4513
Charge Imbalance Calculated, %			2.7	0.6					4.2	-2.6
Acceptable, %			5.1	4.8					4.9	5.3
Relative Saturation										
Gypsum	1.03	1.07	1.20	1.04	1.24	1.33	1.21	1.12	0.92	1.02
CaSO ₃ •0.5H ₂ O	1.43	0.08	1.38	0.00	1.44	0.01	2.06	0.01	1.75	0.01
CaCO ₃	0.05	0.00	0.03	0.00	0.02	0.00	0.13	0.00	0.02	0.00

Table B-4

(Continued)

Description:	10-U-2	10-L-2	11-U-2	11-L-2	12-U-2	12-L-2	13-U-2	13-L-2	14-U-2	14-L-2
Date	11/17/92	11/17/92	11/17/92	11/17/92	11/18/92	11/18/92	11/18/92	11/18/92	11/19/92	11/19/92
Time	1020	1000	1430	1415	1110	1050	1600	1600	1100	1045
Ca, mm/L	23.3	37.3	23.5	36.6	23.8	31.6	23.0	35.2	25.0	34.6
Mg, mm/L			69.1	144.3	54.8	119.2			60.8	151.3
Na, mm/L			18.8	52.6	17.3	41.0			18.7	50.2
Cl, mm/L			85.4	260.8	69.6	232.0			92.9	308.3
CO ₃ , mm/L	3.59	2.11	3.54	2.61	3.48	2.25	3.63	2.45	3.67	2.22
SO ₃ , mm/L	2.15	2.21	1.81	1.79	1.08	0.90	2.23	2.67	0.44	1.22
SO ₄ , mm/L	42.0	59.3	43.3	55.8	38.6	51.9	37.7	54.8	37.8	57.2
Total Hyd SO ₄ , mm/L	52.1	77.9	57.1	72.6	50.9	66.8	45.7	74.3	49.4	77.1
S/N, mm/L	8.0	16.4	11.9	15.0	11.2	14.1	5.7	16.8	11.2	18.7
pH	5.86	3.51	5.8	3.67	6.1	3.77	6.04	3.72	5.95	3.6
Temperature, °C	55	54.4	55.5	54.8	53.8	53.4	56.3	54.7	54	53.5
Ca, mg/L	932	1493	943	1469	956	1267	923	1411	1002	1386
Mg, mg/L			1679	3507	1332	2899			1477	3678
Na, mg/L			433	1210	397	941			431	1155
Cl, mg/L			3029	9245	2469	8226			3294	10929
CO ₃ , mg/L	215	127	212	157	209	135	218	147	220	133
SO ₃ , mg/L	172	177	145	143	86	72	179	214	35	98
SO ₄ , mg/L	4031	5697	4163	5364	3708	4981	3622	5264	3631	5494
Charge Imbalance										
Calculated, %			5.1	3.4	5.0	-1.0			2.8	-2.2
Acceptable, %			5.1	5.3	5.0	5.3			5.0	5.4
Relative Saturation										
Gypsum	1.06	1.28	1.09	1.19	1.14	1.12	1.08	1.27	1.10	1.12
CaSO ₄ •0.5H ₂ O	1.16	0.01	0.91	0.01	0.85	0.01	1.64	0.02	0.30	0.01
CaCO ₃	0.02	0.00	0.01	0.00	0.04	0.00	0.04	0.00	0.03	0.00

Table B-4

(Continued)

Description:	15-U-2	15-L-2	LT-1-UB	LT-1-LB	LT-4-UB	LT-4-LB	LT-5-UB	LT-5-LB
Date	11/19/92	11/19/92	11/21/92	11/21/92	11/24/92	11/24/92	11/25/92	11/25/92
Time	1645	1630	1040	1040	1340	1340	520	520
Ca, mm/L	25.8	37.6	22.2	34.6	22.8	36.5	23.3	35.8
Mg, mm/L			52.7	155.3	60.2	186.7	62.3	189.8
Na, mm/L			23.9	51.3	21.0	62.7	20.3	61.3
Cl, mm/L			106.3	324.0	102.6	405.5	107.1	451.9
CO ₃ , mm/L	3.89	2.58	3.48	2.60	3.79	3.17	1.46	1.48
SO ₃ , mm/L	1.98	2.27	3.32	7.43	3.18	5.96	3.02	5.45
SO ₄ , mm/L	38.1	61.6	39.2	52.7	39.4	55.5	43.9	56.2
Total Hyd SO ₄ , mm/L	52.2	77.5	44.5	69.4	48.2	72.7	49.7	80.0
S/N, mm/L	12.1	13.6	1.9	9.2	5.6	11.2	2.8	18.3
pH	5.99	3.56	5.52	2.58	5.5	3.65	5.74	3.7
Temperature, °C	56.1	55.7	55	54.5	57.6	57	48.3	45.3
Ca, mg/L	1033	1506	889	1388	913	1463	934	1437
Mg, mg/L			1282	3774	1463	4539	1515	4614
Na, mg/L			550	1180	483	1442	467	1408
Cl, mg/L			3770	11486	3638	14376	3795	16018
CO ₃ , mg/L	233	155	209	156	227	190	88	89
SO ₃ , mg/L	158	181	266	595	255	477	242	436
SO ₄ , mg/L	3662	5918	3767	5061	3787	5330	4214	5403
Charge Imbalance Calculated, % Acceptable, %			-3.6 4.9	-0.9 5.5	0.0 5.0	-1.8 5.6	-1.6 5.0	-6.4 5.6
Relative Saturation Gypsum CaSO ₃ •0.5H ₂ O CaCO ₃	1.11 1.43 0.03	1.28 0.01 0.00	1.04 1.09 0.00	1.03 0.04 0.00	1.02 1.02 0.00	0.99 0.04 0.00	1.13 1.33 0.00	1.00 0.03 0.00

Table B-5

Parametric DBA Results

Description:	1-U-1	1-L-1	1-U-3	1-L-3	1-U-4	1-L-4	1-U-6	1-L-6	1-U-7	1-L-7
Adipic, mM	0.128	0.167	<0.035	0.254	0.108	0.567	0.088	0.119	0.179	0.608
Glutaric, mM	0.354	2.400	0.412	0.777	0.253	2.110	0.162	0.674	0.431	1.870
Succinic, mM	0.230	0.865	<0.040	0.247	0.174	0.659	0.089	0.121	0.261	0.817
DBA, mg/L	93	443	47	169	70	439	45	121	114	432

Description:	2-U-1	2-L-1	2-U-3	2-L-3	3-U-2	3-L-2	4-U-2	4-L-2	5-U-1	5-L-1
Adipic, mM	0.154	0.417	0.117	0.350	<0.160	0.481	0.164	0.389	0.294	1.540
Glutaric, mM	0.299	1.620	0.234	0.328	0.128	1.550	0.261	1.660	0.913	5.260
Succinic, mM	0.223	0.533	0.181	0.462	0.067	0.644	0.209	0.367	0.432	2.000
DBA, mg/L	88	338	69	149	25	351	83	319	214	1155

Description:	5-U-3	5-L-3	Tanker DBA-1	5-U-4	5-L-4	5-U-6	5-L-6	6-U-1	6-L-1	6-U-3
Adipic, mM	0.238	1.120	389	0.196	0.566	0.086	0.378	0.249	0.468	0.234
Glutaric, mM	1.080	4.300	1162	1.100	2.580	0.814	1.890	0.768	1.980	0.648
Succinic, mM	0.319	1.560	524	0.478	0.728	0.313	0.521	0.328	0.617	0.171
DBA, mg/L	215	915	272010	230	509	157	366	176	402	140

Description:	Adipic, mM	Glutaric, mM	Succinic, mM	DBA, mg/L
6-L-3	0.495	1.890	0.723	407
7-U-2	0.049	0.627	0.229	117
7-L-2	0.679	2.370	1.020	532
8-U-2	0.185	0.490	0.239	120
8-L-2	0.442	1.540	0.670	347
9-U-1	0.422	1.150	0.539	277
9-L-1	1.650	5.160	2.420	1208
9-U-3	0.322	0.911	0.391	213
9-L-3	2.010	6.660	3.070	1535
9-U-4	0.436	1.610	0.755	365

Description:	Adipic, mM	Glutaric, mM	Succinic, mM	DBA, mg/L
9-L-4	1.160	3.720	1.550	843
9-U-6	0.407	1.320	0.642	309
9-L-6	1.420	4.350	1.730	986
10-U-2	0.594	1.720	0.875	417
10-L-2	1.750	5.440	2.060	1217
11-U-2	0.605	1.840	0.930	441
11-L-2	1.440	4.640	1.700	1023
12-U-2	0.646	1.950	1.000	470
12-L-2	1.640	5.180	2.110	1172
13-U-2	0.550	1.820	0.866	423

Description:	Adipic, mM	Glutaric, mM	Succinic, mM	DBA, mg/L
13-L-2	1.700	5.230	2.050	1180
14-U-2	1.310	3.780	1.840	907
14-L-2	3.090	9.580	3.570	2137
15-U-2	1.260	3.620	1.860	881
15-L-2	2.540	7.950	3.220	1800
Tanker 1	400	1133	546	272384
Tanker 2	11-17-92	11-17-92	11-17-92	260862

(Continued)

Table B-5

Table B-6

**Trace Species
Analytical Data Summary (Parametric Tests)
(Results are in mg/L.)**

Description:	1-U-1		1-L-1		2-U-3		2-L-3	
Element	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit
Aluminum	1.2	0.48	0.52	0.42	0.81	0.42	0.070	0.44
Antimony	0.75	0.24	1.2	0.21	0.83	0.21	1.3	0.22
Arsenic	ND	0.71	ND	0.62	ND	0.63	ND	0.65
Barium	0.43	0.024	0.63	0.021	0.36	0.021	0.74	0.022
Beryllium	ND	0.0048	ND	0.0042	ND	0.0042	ND	0.0044
Boron	400	5.7	1000	12	660	5.1	1200	13
Cadmium	0.17	0.012	0.29	0.010	0.12	0.011	0.32	0.011
Calcium	1100	2.4	1600	2.1	930	2.1	1600	2.2
Chromium	ND	0.024	0.40	0.021	ND	0.021	0.53	0.022
Cobalt	0.095	0.024	0.15	0.021	0.088	0.021	0.20	0.022
Copper	0.050	0.048	0.31	0.042	ND	0.042	0.24	0.044
Iron	ND	0.12	3.9	0.10	ND	0.11	8.4	0.11
Lead	ND	0.12	ND	1.0	0.13	0.11	ND	1.1
Magnesium	1800	2.4	2900	21	1600	2.1	3400	22
Manganese	9.4	0.024	16	0.21	7.8	0.021	19	0.22
Molybdenum	0.16	0.12	0.14	0.10	0.17	0.11	0.15	0.11
Nickel	0.53	0.048	0.95	0.042	0.46	0.042	1.1	0.044
Potassium	49	7.1	93	6.2	41	6.3	110	6.5
Selenium	ND	0.71	1.5	0.62	ND	0.63	1.4	0.65
Silicon	93	2.4	120	2.1	81	2.1	130	2.2
Silver	ND	0.024	ND	0.021	ND	0.021	ND	0.022
Sodium	510	2.4	980	2.1	430	2.1	1100	2.2
Strontium	5.4	0.0071	11	0.0062	5.2	0.0063	11	0.0065
Thallium	0.75	0.24	0.74	0.21	0.35	0.21	0.69	0.22
Vanadium	0.21	0.048	0.46	0.42	0.22	0.042	0.67	0.44
Zinc	0.27	0.048	0.58	0.42	0.19	0.042	0.70	0.44

Description:	5-U-1		5-L-1		9-U-1		9-L-1	
Element	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit	Result	Detect Limit
Aluminum	0.85	0.42	0.76	0.41	0.52	0.44	0.52	0.45
Antimony	0.71	0.21	2.7	2.0	0.23	0.22	0.53	0.23
Arsenic	ND	0.64	ND	0.61	ND	0.66	ND	0.68
Barium	0.35	0.021	0.85	0.020	0.15	0.022	0.25	0.023
Beryllium	ND	0.0042	ND	0.0041	ND	0.0044	ND	0.0045
Boron	330	2.5	1500	12	49	1.3	180	1.4
Cadmium	0.11	0.011	0.36	0.010	0.030	0.011	0.068	0.011
Calcium	900	2.1	1800	20	790	2.2	860	2.3
Chromium	ND	0.021	0.77	0.020	ND	0.022	0.49	0.023
Cobalt	0.098	0.021	0.28	0.020	0.031	0.022	0.048	0.023
Copper	0.045	0.042	0.55	0.041	0.048	0.044	0.59	0.045
Iron	0.14	0.11	9.5	0.10	ND	0.11	5.5	0.11
Lead	0.12	0.11	1.1	1.0	ND	0.11	ND	0.11
Magnesium	1600	2.1	4900	20	370	2.2	770	2.3
Manganese	7.7	0.021	25	0.20	1.8	0.022	3.7	0.023
Molybdenum	0.17	0.11	0.22	0.10	ND	0.11	ND	0.11
Nickel	0.48	0.042	1.4	0.041	0.14	0.044	0.33	0.045
Potassium	40	6.4	130	6.1	17	6.6	30	6.8
Selenium	ND	0.64	1.6	0.61	ND	0.66	1.4	0.68
Silicon	80	2.1	140	2.0	67	2.2	80	2.3
Silver	ND	0.021	ND	0.020	ND	0.022	ND	0.023
Sodium	420	2.1	1300	2.0	160	2.2	290	2.3
Sroutium	4.9	0.0064	12	0.0061	3.8	0.0066	5.7	0.0068
Thallium	0.52	0.21	ND	2.0	0.37	0.22	0.48	0.23
Vanadium	0.23	0.042	0.94	0.41	0.19	0.044	0.57	0.045
Zinc	0.21	0.042	0.96	0.41	0.089	0.044	0.21	0.045

(Continued)

Table B-6

Table B-6

(Continued)

Description:	6-U-3		6-L-3		Method Blank	
Element	Result	Defect Limit	Result	Defect Limit	Result	Defect Limit
Aluminum	0.67	0.40	0.77	0.41	ND	0.20
Antimony	0.72	0.20	2.6	2.1	ND	0.10
Arsenic	ND	0.60	ND	0.62	ND	0.30
Barium	0.30	0.020	0.59	0.021	ND	0.010
Beryllium	ND	0.0040	ND	0.0041	ND	0.0020
Boron	310	2.4	1200	12	ND	0.60
Cadmium	0.11	0.0099	0.29	0.010	ND	0.0050
Calcium	860	2.0	1400	21	ND	1.0
Chromium	ND	0.020	0.71	0.021	ND	0.010
Cobalt	0.096	0.020	0.23	0.021	ND	0.010
Copper	ND	0.040	0.29	0.041	ND	0.020
Iron	ND	0.099	9.7	0.10	ND	0.050
Lead	ND	0.099	1.1	1.0	ND	0.050
Magnesium	1600	2.0	4200	21	ND	1.0
Manganese	7.3	0.020	21	0.21	ND	0.010
Molybdenum	0.18	0.099	0.20	0.10	ND	0.050
Nickel	0.47	0.040	1.2	0.041	ND	0.020
Potassium	40	6.0	110	6.2	ND	3.0
Selenium	1.7	0.60	3.0	0.62	ND	0.30
Silicon	70	2.0	120	2.1	ND	1.0
Silver	ND	0.020	ND	0.021	ND	0.010
Sodium	410	2.0	1200	21	ND	1.0
Strontium	4.5	0.0060	9.4	0.0062	ND	0.0030
Thallium	0.31	0.20	2.6	2.1	ND	0.10
Vanadium	0.23	0.040	0.94	0.41	ND	0.20
Zinc	0.18	0.040	0.68	0.41	ND	0.20

APPENDIX C

Results for the Additive Consumption Tests

Table C-1

Results of Laboratory DBA Analyses for Consumption Test Samples

Description:	LT 0-ASST	LT 1-ASST	LT 1-BSST	LT 1-FT	LT 1-UB	LT 1-UC	LT 1-UD	LT 1-LB
Adipic, mM	0.17	0.08	0.58	0.09	0.38	0.51	0.56	0.85
Glutaric, mM	3.74	3.81	2.59	1.61	1.59	1.49	1.52	3.29
Succinic, mM	<0.08	<0.08	<0.08	0.09	0.95	0.58	0.43	1.14
Adipic, ppmw	24.8	11.1	85.2	12.9	54.8	74.8	81.4	124
Glutaric, ppmw	494	503	342	213	210	197	201	435
Succinic, ppmw	<10	<9	<10	10.5	112	68.1	51.3	135
DBA, ppmw	519	514	427	236	377	340	334	693

Description:	LT 1-LC	LT 1-LD	LT 1-RWT	LT 1-BQBT	LT 2-LD	LT 2-RWT	LT 3-RWT	LT 4-ASST
Adipic, mM	0.75	0.81	0.68	0.76	0.77	0.57	0.54	<0.06
Glutaric, mM	2.74	3.05	2.62	3.16	3.50	2.72	2.70	3.92
Succinic, mM	0.73	0.81	0.60	0.79	0.89	0.61	0.66	<0.07
Adipic, ppmw	110	118	99	111	113	83	79	<8
Glutaric, ppmw	362	403	346	418	462	359	357	518
Succinic, ppmw	85.6	95.1	71.2	93.8	105	72.5	77.9	<8
DBA, ppmw	557	616	516	623	680	514	514	518

Table C-1

(Continued)

Description:	LT 4-BSST	LT 4-FT	LT 4-UB	LT 4-UC	LT 4-UD	LT 4-LB	LT 4-LC
Adipic, mM	0.07	<0.06	0.51	0.491	0.49	0.72	0.74
Glutaric, mM	2.86	1.87	1.66	1.43	1.68	3.59	3.42
Succinic, mM	<0.08	0.11	0.81	0.57	0.57	0.88	0.96
Adipic, ppmw	<9	<9	73.9	71.9	71.0	105	108
Glutaric, ppmw	378	247	219	189	222	474	452
Succinic, ppmw	<9	13.1	95.3	67.0	67.0	104	113
DBA, ppmw	389	260	389	328	360	683	673

Description:	LT 4-LD	LT 4-RWT	LT 4-BQBT	LT 5-ASST	LT 5-BSST	LT 5-FT	LT 5-UB
Adipic, mM	0.582	0.488	0.535	<0.058	0.075	<0.063	0.158
Glutaric, mM	3.48	2.96	3.09	3.85	2.88	1.91	1.34
Succinic, mM	0.751	0.592	0.732	<0.072	0.350	0.064	0.554
Adipic, ppmw	85.1	71.3	78.2	<8	11.0	<9	23.1
Glutaric, ppmw	459.7	391.0	408.2	508.6	380.5	252.3	177.0
Succinic, ppmw	88.7	69.9	86.4	<8	41.3	7.6	65.4
DBA, ppmw	633.5	532.3	572.8	508.6	432.8	259.9	265.5

Table C-1

(Continued)

Description:	LT 5-UC	LT 5-UD	LT 5-LB	LT 5-LD	LT 5-RWT	LT 5-BQBT	LT-5 DBA Tanker
Adipic, mM	0.20	0.27	0.52	0.45	0.36	0.48	398
Glutaric, mM	1.36	1.56	3.46	3.41	3.02	3.36	1,167
Succinic, mM	0.34	<0.07	0.79	0.60	0.50	0.64	539
Adipic, ppmw	29.8	39.8	76.0	66.3	51.9	69.6	54,300
Glutaric, ppmw	180	206	457	451	399	444	144,000
Succinic, ppmw	40.5	<9	93.5	70.9	58.6	75.9	59,400
DBA, ppmw	250	246	627	588	509	589	258,000

Values below the detection limit are considered to be 0.

Vessel Key:

ASST = "A" Sludge Surge Tank
 BSST = "B" Sludge Surge Tank
 BQBT = "B" Quencher Blowdown Tank
 FT = Filtrate Tank
 RWT = Return Water Tank
 U(B,C,D) = Module B, C, or D Absorber Feed Tank
 L(B,C,D) = Module B, C, or D Quencher

Table C-2

DBA Consumption Test Results

Period	Average Ad-dition Rate (lb/hr)	Average Blowdown Rate (lb/hr)	Average Inventory Change Rate (lb/hr)	Average Nonsolubion Loss Rate (lb/hr)	Average Nonsolubion Loss Rate (lb/ton of SO ₂ removed)	Average AFT DBA Con- centration (ppmw)	Average SO ₂ Removal Effi- ciency (%)	Average Unit Load (MW)
Succinic Acid								
Inv 1 - Inv 4 (75.0 hrs)	12.1	1.7	-0.01	10.4	1.1	77	95.5	407
Inv 4 - Inv 5 (15.4 hrs)	8.4	0.5	-12.2	20.1	2.3	56	95.3	383
Inv 1 - Inv 5 (90.4 hrs)	11.4	1.5	-2.1	12.0	1.3	63 ± 24	95.3	403
Glutaric Acid								
Inv 1 - Inv 4 (75.0 hrs)	31.9	7.2	4.2	20.5	2.2	206	95.5	407
Inv 4 - Inv 5 (15.4 hrs)	22.4	3.0	-7.6	27.0	3.1	199	95.3	383
Inv 1 - Inv 5 (90.4 hrs)	30.3	6.5	2.2	21.6	2.4	200 ± 12	95.3	403
Adipic Acid								
Inv 1 - Inv 4 (75.0 hrs)	13.2	1.4	-1.7	13.5	1.5	71	95.5	407
Inv 4 - Inv 5 (15.4 hrs)	9.3	0.5	-13.8	22.6	2.6	52	95.3	383
Inv 1 - Inv 5 (90.4 hrs)	12.5	1.2	-3.7	15.0	1.7	58 ± 17	95.3	403
Total DBA								
Inv 1 - Inv 4 (75.0 hrs)	57.2	10.3	2.5	44.4	4.8	354	95.5	407
Inv 4 - Inv 5 (15.4 hrs)	40.1	4.0	-33.4	69.7	8.0	306	95.3	383
Inv 1 - Inv 5 (90.4 hrs)	54.2	9.2	-3.7	48.7 ± 7	5.4 ± 1	321 ± 42	95.3	403

Table C-3**Assumptions Used in DBA Consumption Uncertainty Analysis**

Variable	Estimated Error	Notes
DBA Concentrations	10%	
Tank Volumes	10%	
Totalizer Volume (change)	10%	
SO ₂ Outlet Concentrations	20%	
SO ₂ Inlet Concentration	0.13 lb/MMBtu	Standard Deviation
Coal Heating Value	66 Btu/lb	Standard Deviation
Stack Gas Flow Rate	20%	
DBA Tanker Level	1/4" (35 gal)	
Unit Load	5%	
Unit Heat Rate	5%	
DBA Specific Gravity	0.0075	Standard Deviation
Slurry Liquid Volume Fraction	0	Simplification

APPENDIX D

Results of Solids Settling Properties Tests

Table D-1

Settling Test Results

Test No.	Date	Time	Average DBA Concentration (ppm)		UA @ 30 wt. % (ft ³ -day/ton) Initial Wt. % Solids in Parentheses	UA @ 30 wt. % (ft ³ -day/ton) Initial Solids Corrected to 17.2%	Final UF Solids (wt. %)	24 Hours Later (ft ³ -day/ton) Initial Solids Corrected to 17.2%	Final UF Solids 24 Hrs Later (wt. %)
			Quench	AFT					
Baseline	09-29-92	1300-1700	0	0	0.8 (17.2%)	0.8	64.7		
Parametric Test 1	11-05-92	0900	330	73	1.0 (17.9%)	1.1	56.2		
Parametric Test 2	11-06-92	0800	240	78	1.2 (17.8%)	1.3	55.3	1.9	55.1
Parametric Test 5	11-12-92	1400	740	200	0.7 (22.8%)	1.7	52.7		
Parametric Test 6	11-13-92	1400	410	160	0.8 (18.1%)	0.9	55.4	1.4	54.6
Parametric Test 9	11-16-92	1500	1140	290	2.8 (19.6%)	3.8	44.7		
Parametric Test 11	11-17-92	1530	1020	440	3.0 (19.4%)	4.1	43.3	7.3	43.2

Table D-2

Filter Leaf Test Conditions and Results

Summary of Test Conditions				
Filter Cloth Type		POPR 873		
Cloth Area		0.0873 ft²		
Form Filtration Vacuum		20.00 in. Hg		
Cake Solids Drying Time		120 sec		
Cake Solids Test Vacuum		14 - 19.5 in. Hg		
Cake Thickness		0.88 - 1.00 in.		
Test Results				
Test	Date	Time	Final Cake Sol- ids Wt. %	Form Fil- tration Rate (lb/hr/ft²)
Baseline	09-29-92	1300-1700	78.3	390
Parametric Test 1	11-05-92	0900	79.9	650
Parametric Test 2	11-06-92	0800	85.8	730
Parametric Test 5	11-12-92	1400	83.2	600
Parametric Test 6	11-13-92	1400	85.7	910
Parametric Test 9	11-16-92	1500	76.6	320
Parametric Test 11	11-17-92	1530	78.4	520
Long Term Test Inventory 1	11-21-92	1030	83.3	1200
Long Term Test Inventory 4	11-24-92	1330	82.4	790
Long Term Test Inventory 5	11-25-92	0530	82.5	920

APPENDIX E

Details of FGDPRISM Calibration for Big Bend Unit 4

E.1 MODEL DESCRIPTION

FGDPRISM uses the two-film theory to predict inter-phase mass transfer. The theory is based on the assumption that the transfer of a species can be modeled as diffusion through thin, stagnant films that exist on both the gas and liquid sides of the gas-liquid interface. It also assumes that the liquid and gas are at equilibrium at the interface and that the interface provides no resistance to mass transfer. Early versions of the FGDPRISM model used an average, constant value for the liquid-side film thickness throughout the spray section of an absorber. A more rigorous approach was instituted with Version 2.0 wherein the liquid-side film thickness increases as the droplets travel through the absorber. This is a more accurate representation because, as the droplets exit a spray nozzle, they have a great deal of internal turbulence which promotes mass transfer (i.e., a thinner liquid film). As the droplets travel further down the tower, this turbulence decays and the mass transfer rate decreases. Therefore, in Version 2.0 an algorithm was implemented to change the liquid-film thickness as a function of droplet residence time. This modification resulted in less sensitivity to tower height or spacing between headers than was predicted with earlier versions.

Another important modification included in Version 2.0 is the limestone dissolution methodology. Previously, the model computed the contribution that solid-phase CaCO_3 made to SO_2 removal because of dissolution within the absorber by using a simple dissolution rate equation, where only a single rate constant was required. The methodology used in Version 2.0 is a combined surface reaction/diffusion rate model, where the limestone dissolution rate is controlled by two series resistances:

- 1) Diffusion of chemical species through a stagnant "film" surrounding the dissolving limestone particle; and
- 2) A surface reaction rate that is calculated from an empirical equation which accounts for the inhibiting effects of species such as sulfite and magnesium.

The diffusion rate is a function of the film thickness and the concentrations of species such as calcium and carbonate at the limestone surface and in the bulk solution. Values for the limestone diffusion film thickness in the absorber spray and the reaction tank are calculated by the model.

The surface reaction rate is a function of the solution composition at the limestone surface and the limestone reactivity. The rate is calculated from the following equation:

$$\text{Surface Reaction Rate} = \frac{k \cdot (1 - \text{CaCO}_3 \text{ RS})^{3.0} \cdot A}{(\text{CaCO}_3 \text{ RS}') \cdot (\text{CaSO}_3 \text{ RS})} \quad (\text{E-1})$$

where: k = rate constant;
 $\text{CaSO}_3 \text{ RS}$ = relative saturation of CaSO_3 ;
 $\text{CaCO}_3 \text{ RS}'$ = $\text{CaCO}_3 \text{ RS}$ corrected for the effect of magnesium, that is:
 $\text{CaCO}_3 \text{ RS}' = \text{CaCO}_3 \text{ RS} + (576.13 \cdot \text{MgCO}_3 \text{ RS})$; and
 A = limestone surface area.

The exponent 3.0 in the surface reaction rate equation and the constant 576.13 in the corrected CaCO_3 relative saturation expression are based on experimental data from EPRI's HSTC. No adjustments are typically made to these values as part of the calibration procedure. The surface area is calculated by the model from the input limestone feed particle size distribution, and can be adjusted using an empirical multiplier called the surface area factor.

As part of the model calibration procedure, the overall limestone dissolution rate is adjusted by changing the surface area factor and the limestone reaction rate constant (k) to match the observed limestone utilization as a function of pH in the reaction tank.

The calibration procedure also impacts gas-liquid mass transfer rate calculations, with both gas-film thicknesses and liquid-film thicknesses being adjusted to match the mass

transfer characteristics of the particular absorber. For the spray sections of an absorber, the model predicts gas/liquid surface area by determining the trajectory of each slurry droplet as it passes through the absorber and calculating its residence time. These calculations cannot be verified, however, since there is no method of directly measuring the surface area of the spray in an absorber. In addition, the model is unable to predict the gas/liquid distribution in a tower, which can greatly affect scrubber performance. To address these problems, the mass transfer film coefficients are varied to match observed SO₂ removals, since the surface area and film coefficients together determine the SO₂ removal performance. For the Big Bend Unit 4 absorber, separate film thicknesses are specified for the packing and spray sections of the absorber loop, and for the spray section of the quench loop.

E.2 BIG BEND UNIT 4 CALIBRATION RESULTS

To calibrate FGDPRISM Version 2.0 for the Big Bend Unit 4 FGD system, the limestone reaction rate constant and surface area factor and the liquid- and gas-film thicknesses were adjusted for each section of the absorber to give the best overall match to measured pH and SO₂ removal values. The results for four baseline tests and one parametric test at a relatively high DBA concentration were used for this calibration. The parametric test was included to ensure that the calibration would be accurate at high SO₂ removal and with DBA added.

The reaction rate constant and surface area factor for each loop were adjusted to achieve the observed split between the absorber- and quench-loop pH values. The limestone reaction rate constants used were 7.5×10^{-6} for the absorber loop and 9.0×10^{-9} for the quench loop. The absorber- and quench-loop surface area factors were 1.0 and 0.1, respectively. The low reaction rate constant and the low surface area factor were required in the quench loop to predict the very low pH values observed.

The liquid- and gas-film thicknesses were adjusted to match the measured absorber- and quench-loop SO₂ removals. Because most of the removal occurs in the absorber loop, primarily in the packing, the model's predictions are more sensitive to changes in the

absorber-loop film thicknesses. Therefore, more emphasis was placed on the absorber-loop film thicknesses for the calibration.

The liquid-film thicknesses for the spray and droplet rain sections were fixed at 0.1% of the droplet diameter. The liquid-film thickness of the packing was then adjusted to match the removal data. The final value arrived at for the best fit was 0.017% of the "droplet diameter" for the packing. The packing droplet diameter actually is a term used to determine the liquid holdup in the packing. Larger diameters result in higher holdup amounts for a fixed surface area, since volume is proportional to the area times the diameter.

These droplet diameter percentages correspond to film thicknesses of 5.5 microns for the spray header that floods the packing, 1.0 micron for the packing, 5.0 microns for the droplets that rain from the packing, and 1.28 microns for the spray headers below the packing.

The gas-film thickness was assumed to be the same for all sections of the absorber loop, at 56 microns. This thickness value was based on results from the high DBA test conditions where SO_2 removal was essentially gas-film limited; there is little or no liquid-side resistance at these conditions.

In the quench loop, 0.1% of the droplet diameter was used for the liquid film thickness, which corresponds to 1.9 microns. In the five tests used for the calibration, there were extremely low pH values in the quench loop, but higher than expected SO_2 removals of 10% to 20% there. Consequently, the gas-film thickness in the quench loop was reduced to 10 microns, from the 56-micron value of the absorber loop, to allow these moderate removal levels to be predicted with little alkalinity in the scrubbing liquor.

APPENDIX F
Supporting Information for the Economic Analyses

F.1 COST BASES

In Section 4 of the report, Table 4-2 summarized values for the various economic factors used in the economic evaluations. The development of these factors is discussed below.

The installed capital cost of a DBA additive system for Big Bend Unit 4 was estimated by Stone & Webster Engineering Corporation (SWEC). Budgetary quotes for delivered cost of the DBA additive were provided by Monsanto and DuPont. Other costs for the first upgrade option considered in Section 5 (DBA addition with two feet of packing and two upper-loop slurry recirculation pumps operating) include additional limestone reagent, operating and maintenance charges for increased gypsum production, and offsetting revenues for increased gypsum sales. Most of these cost data were provided by Teco.

The installed capital cost for a DBA additive system at Big Bend Unit 4, was estimated by SWEC at \$550,000 ($\pm 30\%$) for a design capacity of 100 lb/hr as pure DBA. The system design is based on storage and injection of a 50% aqueous DBA solution. To evaluate this option over a range of potential DBA concentrations, the capital cost of the DBA system was scaled from the 100 lb/hr base case using the simple equation shown in Table 4-2. SWEC has determined that the cost of the additive system is not very sensitive to capacity. Much of the cost is associated with the insulated, heated, stainless steel DBA storage tank. The capacity of this tank is based on a minimum of two tanker truck loads, and the tank size would not be changed with DBA injection rate over this range of injection rates. At a 100 lb/hr DBA feed rate, two tanker truck loads of 50% DBA represent a 19-day supply.

For comparison, the cost of a DBA additive system included in EPRI's FGDCOST spreadsheet model (developed by United Engineers) was calculated based on the same capacity. The SWEC estimate is about 50% higher than the EPRI estimate, but the EPRI system design basis is not described in sufficient detail to evaluate differences in these estimates. The FGDCOST spreadsheet also shows that the cost is not very sensitive to capacity.

Budgetary cost estimates for 50% aqueous DBA delivered to Big Bend were obtained by SWEC from Monsanto in Pensacola, FL and from DuPont in Victoria, TX. Monsanto quoted a delivered price of about \$0.21/lb on a dry DBA basis. The DuPont price was about \$0.33/lb of dry DBA. The DBA upgrade option was evaluated using both of these prices.

TECo provided several of the other pertinent cost factors. They estimated that production of additional gypsum will incur a marginal operating and maintenance charge of \$1.30/ton. No capital improvements are required. TECo's estimated cost for prepared limestone is \$7.80/ton. A limestone consumption rate of 1.6 tons per ton of additional SO₂ removal was used in the calculation. TECo also receives revenue from gypsum sales, but the terms of their sales agreement cannot be disclosed. A conservative gypsum sales price of \$2.50/ton was assumed for the economic evaluation. This price is about 20% below the lowest reported cost for naturally produced gypsum.

Options involving changes in the packing depth used in the absorber section will change the annual maintenance charge for replacing packing. TECo has estimated that replacing 1 foot of packing costs \$25,000/yr per module for the first foot of packing. Replacing the second and third feet of packing are estimated to cost \$21,000 and \$17,000/yr per module, respectively.

Changing the packing depth will also change the module pressure drop and fan power. TECo provided packing pressure drop estimates from a vendor proposal for the Big Bend Unit 3 FGD system, which would use a similar packing. The value provided corresponds to 1 inch per foot at a flue gas velocity of 7.5 ft/s. Therefore a pressure drop of 1.0 in. H₂O per foot of packing has been assumed for this evaluation. The sensitivity of upgrade economics to the packing pressure drop was discussed in Section 4.3.

The effect of additional module pressure drop on fan power consumption was calculated assuming that fan efficiency does not change with inlet vane position. In practice, additional module pressure drop would be accommodated by opening the fan inlet vanes to maintain the same flue gas flow rate. Typically, overall fan efficiency would be expected to

increase somewhat as the inlet vanes are opened. The assumption that fan efficiency is constant would tend to overstate the cost of additional module pressure drop. This assumption would be somewhat conservative for cases with increased module pressure drop, but slightly optimistic for options with decreased module pressure drop. The cost of 1 in. H₂O of additional fan pressure drop was calculated to be \$66,000/yr for the entire FGD system, based on 70% overall fan/motor efficiency and a 65% Unit 4 capacity factor.

To evaluate changes in SO₂ removal efficiency for options that involve changing the number of upper-loop recycle pumps operating (one or three versus the normal two per module) with FGDP_{PRISM}, assumptions had to be made regarding the slurry flow rates for each situation. Changing the number of pumps in service will also change fan power and pump power costs. Pump power consumption was measured during baseline and parametric tests. With two pumps operating, the average power consumption was 287 kW per pump. The corresponding annual cost (@ \$0.05/kWh) for 8000 hours of operating time is \$115,000 per pump.

For options with one or three pumps operating, it was assumed that after nozzle or header changes, each pump motor would operate at the same power consumption. In this application, the pump motor power consumption will be proportional to slurry flow rate and pump discharge pressure. The discharge pressure consists of three components: nozzle pressure requirements, static head requirements, and slurry header pressure drop. It was assumed that with nozzle changes, the new nozzles would be sized to operate at the current slurry feed pressure. The static head requirements also would not change. However, the pressure drop through the slurry headers would be affected by the slurry velocity, i.e., the number of pumps operating. The pressure drop would be lower for the one-pump case, but higher for the three-pump case.

In the one-pump case, this would mean that, at constant motor power consumption, the slurry flow would be greater than one-half the two-pump flow rate. Correspondingly, the three-pump flow rate would be less than 50% greater than the two-pump flow rate. It would be possible to calculate new slurry flow rate values for the one-pump and three-pump cases from pump performance curves, friction loss calculations for the absorber piping and header configura-

tions, and pressure/flow characteristics of candidate replacement nozzles. However, this level of detail did not seem warranted for this evaluation. Based on the judgment of the evaluators, it was assumed that the one-pump case would produce 60% of the slurry flow of the normal two pump case at 50% of the power consumption, and the three-pump case would produce 140% of the two-pump flow rate at 150% of the power consumption.

The one-pump case was therefore modeled with FGDPRISM assuming that pump would deliver 60% of the normal two-pump flow. The three-pump case was not simulated with FGDPRISM. Based on full-scale tests results and the one-pump model results, it was assumed that NTU would increase by 40% for the three-pump case compared to the two-pump case values at similar upper-loop DBA concentrations.

Changing the number of operating pumps will also change the module pressure drop. The relationship between module pressure drop and upper-loop slurry flow rate was not measured during the tests at normal flue gas velocity, but can be estimated from the measured change in module pressure drop for two- versus three-pump operation at high gas velocity.

When the unit was operating with two pumps in service in the absorber loop, and with a superficial gas velocity of 7.5 ft/sec, the module pressure drop was measured at about 4.7 in. H₂O. Raising the flue gas velocity to 11.1 ft/sec raised the module pressure drop to 6.2 in. H₂O. Turning on the third pump at these high-velocity conditions raised the pressure drop further to 7.3 in. H₂O, but also decreased the flue gas superficial velocity to 9.4 ft/sec. Slurry velocity measurements indicated that operating with the third pump in service raised the slurry flow rate by as much as 30%. Based on these data, it is estimated that operation with three pumps at 140% of the two-pump flow, at the normal flue gas superficial velocity of 7.5 ft/sec, will increase the module pressure drop by 2 in. H₂O. Operation with one pump at 60% of the two-pump flow is estimated to decrease the module pressure drop by a like amount. The corresponding change in fan power cost is 2 x \$66,000/yr, or \$132,000/yr. The sensitivity of the economics to this pressure drop relationship was discussed in Section 4.3 of the report.

F.2 RESULTS

F.2.1 Economics for DBA Addition in the Current FGD Configuration

The economic factors described above were included in spreadsheet calculations to estimate the costs for additional tons of SO₂ removed by the various SO₂ removal efficiency upgrade options. Table F-1 summarizes details of the cases for DBA addition in the current operating configuration (two feet of packing and two operating pumps in the absorber-loop section) based on a delivered price of \$0.21/lb for DBA, and a 65% Unit 4 capacity factor. Much of this information was previously presented in Table 4-3, although not all details were included there. Examples of the calculations involved in developing the entries in the table are included later in this appendix.

The SO₂ removal performance values in Table F-1 are based on the actual parametric test performance data. The expected overall SO₂ removal efficiency is entered in Column 3 for increasing levels of DBA concentration, which are listed in Column 2. Column 4 shows the tons of SO₂ still available for removal on an annual basis, based on 13.2 tons of SO₂ per hour in the Unit 4 FGD system inlet gas at full load, the SO₂ removal efficiency in Column 3, and a 65% Unit 4 capacity factor.

DBA consumption in lb/hr at full load is shown in Column 5. These DBA consumption rates have been scaled from the long-term test results, assuming that consumption is directly proportional to DBA concentration and to SO₂ removal rates. DBA additive system annual capital charges (amortized over 10 years at 11.5%) are shown in Column 6. SWEC's capital cost estimate, based on a design feed rate of 100 lb/hr, has been scaled using the equation shown in Table 4-2. Annual additive costs are calculated in Column 7 based on the full-load feed rate and a Unit 4 capacity factor of 65%.

Columns 8, 9, and 10 show the annual incremental costs (compared to the base case) for limestone, operating and maintenance charges, and annual additional gypsum revenue for each level of SO₂ removal. These costs were estimated from the factors provided by TEGCO.

Column 11 sums the total net annual cost (compared to the base case) of each optional case. This represents the total for Columns 5 through 9, minus the revenue estimate in Column 10.

Column 12 is the calculated marginal cost of additional tons of SO₂ removed, as the DBA concentration is increased from case to case. Note that the first DBA case includes the bulk of the capital investment, while the increased costs for the subsequent cases reflect only the costs associated with enlarging the feed system.

Column 13 also reflects the marginal cost of additional tons of SO₂ removed as the DBA concentration is increased, but calculated in a different manner. The values in Column 12 are referred to the case immediately above in the table, while the values in Column 13 are all referred back to the base case. Thus the values in Column 13 represent an average marginal cost for all of the additional tons of SO₂ removed, when going from baseline performance to the removal efficiency listed in Column 3.

The final two columns are the calculated annual values for additional tons of SO₂ removed. This is defined as the annual value of the additional SO₂ removed minus the annual additional costs that were summed in Column 11. The results in Column 14 assume that emissions credits are valued at \$250/ton, while the results in Column 15 assume they are valued at a more conservative value of only \$150/ton.

F.2.2 Economics for Modeled Upgrades

Two other upgrade options were considered: varying the depth of packing in the upper-loop sections of the modules, and changing the number of operating pumps. These

potential upgrade options were not tested a full scale. SO₂ removal performance results for these options are based on predictions from the calibrated FGDPRISM model.

The costs for these options were also evaluated in spreadsheet calculations. Table F-2 details the cost estimate components for these cases. Some of these values were summarized in Table 4-4, although in less detail. Also included in Table F-2 is the base case performance. This base case is used as a reference point for the marginal costs in the first case for each option, and as the reference point for all of the "average cost" calculations.

The first upgrade option evaluated is adding 1 foot of packing to the upper-loop sections of the modules. Even for this no-DBA case, some costs are associated with adding the additional foot of packing. The capital cost for the third foot of packing is estimated at a total of \$68,000 per year for the four modules. The additional flue gas pressure drop (1 in. H₂O per foot of packing) increases fan power costs by about \$66,000/yr. Table F-2 has additional columns relative to Table F-1 to allow these cost components to be reflected.

The next series of cases in Table F-2 is based on removing 1 foot of packing. The first calculated marginal cost for this series of cases is actually a "savings," as the net cost is reduced, but the tons of SO₂ removed are also reduced. The annual value of this case is negative because the value of the credits at even the lower \$150/ton value far exceeds the \$21/ton saved.

The third upgrade option evaluated in Table F-2 is based on taking one of the two operating upper-loop slurry recycle pumps out of service, with the normal two feet of packing in the modules. FGDPRISM has predicted that NTU is roughly proportional to the upper-loop slurry flow rate, which is consistent with the limited L/G test results obtained by operating the third pump. As described above, it was assumed that after nozzle changes, 60% of the normal, two-pump flow can be obtained with one pump because the frictional losses in the slurry piping will decrease. It was also assumed that no other equipment changes would be required. Table F-2 includes columns for the pump and fan power cost savings, and for an annual capital charge for the nozzle replacement.

The final option evaluated in Table F-2 is operation with three upper-loop slurry recycle pumps. The slurry flow was assumed to be 140% of the two-pump flow, and the SO₂ removal performance was assumed to be 40% greater (in terms of NTU) than the corresponding two-pump cases. Table F-2 includes columns for the additional pump and fan power costs, and for an annual capital charge for the nozzle replacement.

F.3 EXAMPLE CALCULATIONS

This section summarizes some of the calculations used to derive the values presented in Tables F-1 and F-2. Note that in some of the example calculations, the results shown do not agree exactly with the values listed in the table due to rounding errors.

F.3.1 Example Calculations For Table F-1

Change in Total Annual Cost (Column 11) = DBA Capital (Col. 6) + DBA consumed (Col. 7) + Addt'l Reagent (Col. 8) + Addt'l O&M (Col. 9) - Addt'l Gypsum Revenue (Col. 10)

For the first DBA case, this is calculated as: \$78,000 + \$36,000 + \$21,000 + \$6,000 - \$11,000, or \$130,000.

Marginal Cost (Column 12) = (Change in Total Cost (Col. 11) - Change in Total Cost for previous case (Col. 11))/(Additional Tons of SO₂ Available for previous case (Col. 4) - Additional Tons of SO₂ Available for current case (Col. 4))

For the first DBA case, this is calculated as: (\$130,000 - \$0)/(5051 tons - 3386 tons), or \$78/ton. For the second DBA case, the calculation is (\$190,000 - \$130,000)/(3386 tons - 1858 tons), or \$39/ton.

Average Cost (Column 13) = (Change in Total Cost relative to base case (Col. 11)/(Additional Tons of SO₂ Available for base case (Col. 4) - Additional tons of SO₂ Available for current case (Col. 4))

For the first DBA case, the numbers are essentially the same as for the first case above: \$130,000/(5051 tons - 3386 tons), or \$78.1/ton. For the second DBA case, the numbers are: \$190,000/(5051 tons - 1858 tons), or \$59.5/ton.

Net Annual Value (Column 14) = (\$250/ton - Average Cost (Col. 13)) x (Additional Tons of SO₂ Available for base case (Col. 4) - Additional tons of SO₂ Available for current case (Col. 4))

For the first DBA case, the numbers are (\$250/ton - \$78/ton) x (5051 tons - 3386 tons), or \$286,380. For the second DBA case, the numbers are (\$250/ton - \$59/ton) x (5051 tons - 1858 tons), or \$609,863.

Net Annual Value (Column 15) = (\$150/ton - Average Cost (Col. 13)) x (Additional Tons of SO₂ Available for base case (Col. 4) - Additional tons of SO₂ Available for current case (Col. 4))

For the first DBA case, the numbers are (\$150/ton - \$78/ton) x (5051 tons - 3386 tons), or \$119,880. For the second DBA case, the numbers are (\$150/ton - \$59/ton) x (5051 tons - 1858 tons), or \$290,563.

F.3.2 Example Calculations For Table F-2

Change in Total Annual Cost (Column 15). The calculation remains basically the same as for Table F-1 above expect there are additional entries for Addt'l Packing Capital (Col. 7), Addt'l Fan kW (Col. 9), Addt'l Pump kW (Col. 10), and Nozzle Change Capital (Col. 14) that are included in the total. However, the Addt'l Packing Capital numbers only apply to the varied packing depth cases, and the Addt'l Pump kW and Nozzle Change Capital numbers only apply to the cases where the number of pumps operating are varied. For the cases where these number do not

apply, there are zeros in the table for these columns. Also, in some cases the column numbers for the equation terms for Table F-1 change in Table F-2 because of the additional columns.

Marginal Cost (Col. 16), Average Cost (Col. 17), and Net Annual Value (Cols. 18 and 19).

These calculations remain the same as for Table F-1, except some of the column numbers are changed because of the additional columns in Table F-2.

Table F-1

Base Case Economic Calculations
\$0.21/lb of DBA, 65 % Capacity Factor, 1" ΔP/ft of Packing

(1) Packing Depth (ft)	(2) Upper Loop DBA Conc. (ppm)	(3) SO ₂ Removal (%)	(4) Addit'l SO ₂ Tons of Available SO ₂ (per year)	(5) DBA Feed (lb/hr)	(6) DBA Capital (\$/yr)	(7) DBA (\$1000/yr)	(8) Addit'l Re- agent (\$1000/yr)	(9) Addit'l O&M (\$1000/yr)	(10) Addit'l Gypsum Revenue (\$1000/yr)	(11) Change in Total Annual Cost (\$1000/yr)	(12) Marginal Cost (\$/ton SO ₂)	(13) Average Cost (\$/ton SO ₂)	(14) Net Annual Value (\$1000) @ \$250/ton	(15) @ \$150/ton
2	125	95.5	3386	30	78	36	21	6	11	130	78	78	287	120
2	250	97.5	1858	61	87	73	40	11	21	190	39	59	608	289
2	500	99.2	619	125	97	149	55	15	30	287	78	65	822	378
2	750	99.6	307	187	103	224	59	17	32	371	271	78	815	341
2	1000	99.8	169	250	107	299	61	17	33	452	585	93	769	281
Add DBA														
2	0	93.3	5051	0	0	0	0	0	0	0	0	0	0	0
Base Case														

Table F-2

Modeled Case Economic Calculations
\$0.21/lb of DBA, 65% Capacity Factor, 1 in. H₂O/ft of Packing

Packing Depth	Upper Loop DBA Conc., ppm	SO ₂ Removal	Add'l Tons SO ₂ Available (per year)	DBA Feed, lb/hr	DBA Capital, \$1000/yr	Add'l Packing Capital, \$1000/yr	DBA, \$1000/yr	Add'l Fan kW, \$1000/yr	Add'l Pump kW, \$1000/yr	Add'l Reagent, \$1000/yr	Add'l O&M, \$1000/yr	Add'l Gypsum Revenue, \$1000/yr	Nozzle Change Capital, \$1000/yr	Change in Total Annual Cost, \$1000/yr	Marginal Cost, \$/ton SO ₂	Average Cost, \$/ton SO ₂	Net Annual Value (\$1000)	
																	@ \$250/ton	@ \$150/ton
Base Case (2 ft of packing, 2 upper-loop pumps in service, no DBA)																		
2	0	93.3	5051	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
With 3 ft of packing																		
3	0	98.0	1506	0	0	68	0	66	0	44	12	24	0	167	47	47	719	365
3	125	98.8	914	31	78	68	37	66	0	52	14	28	0	288	204	70	747	333
3	250	99.2	637	62	87	68	74	66	0	55	15	30	0	336	176	76	767	326
3	500	99.5	357	125	97	68	149	66	0	59	16	32	0	423	311	90	750	281
3	750	99.7	196	188	103	68	224	66	0	61	17	33	0	506	514	104	708	222
3	1000	99.8	151	250	107	68	299	66	0	61	17	33	0	586	1783	120	639	149
With 1 ft of packing																		
1	0	76.1	17987	0	0	-84	0	-66	0	-161	-45	-87	0	-270	21*	21*	-4122	-1671
1	125	79.2	15637	26	76	-84	31	-66	0	-132	-37	-71	0	-141	55	13*	-3663	-1447
1	250	81.9	13594	53	85	-84	63	-66	0	-107	-30	-57	0	-81	29	10*	-3212	-1200
1	500	86.5	10172	110	95	-84	132	-66	0	-64	-18	-34	0	29	32	-6*	-2467	-798
1	750	89.7	7765	171	101	-84	204	-66	0	-34	-9	-18	0	131	42	-48*	-1966	-538
1	1000	91.9	6108	233	106	-84	278	-66	0	-13	-4	-7	0	225	57	-213*	-1646	-383

(Continued)

Table F-2

Packing Depth	Upper Loop DBA Conc., ppm	SO ₂ Removal	Add'l Tons SO ₂ Available (per year)	DBA Feed, lb/hr	DBA Capital, \$1000/yr	Add'l Packing Capital, \$1000/yr	DBA, \$1000/yr	Add'l Fan kW, \$1000/yr	Add'l Pump kW, \$1000/yr	Add'l Revenue, \$1000/yr	Add'l O&M, \$1000/yr	Add'l Gypsum Revenue, \$1000/yr	Nozzle Change Capital, \$1000/yr	Change in Total Annual Cost, \$1000/yr	Marginal Cost, \$/ton SO ₂	Average Cost, \$/ton SO ₂	Net Annual Value (\$1000)	Net Annual Value @ \$250/ton SO ₂	Net Annual Value @ \$150/ton SO ₂
2	0	80.2	14874	0	0	0	0	-132	-345	-123	-34	-66	16	-552	58*	58*	-1888	-906	
2	250	89.2	8082	57	86	0	68	-132	-345	-38	-11	-20	16	-335	32	111*	-406	-119	
2	500	92.4	5695	117	96	0	140	-132	-345	-8	-2	-4	16	-231	44	359*	86	135	
2	1000	96.1	2914	242	107	0	290	-132	-345	27	7	14	16	-45	67	-21	595	365	
2	2000	98.7	1000	496	119	0	593	-132	-345	51	14	27	16	288	174	71	741	320	
2	2500	99.2	606	623	123	0	745	-132	-345	55	16	30	16	448	405	101	679	219	
With 3 pumps (plus nozzle change); 140% of flow @ 150% of power																			
2	0	97.7	1715	0	0	0	0	132	345	42	12	22	16	524	152	157	326	-24	
2	250	99.4	423	62	87	0	75	132	345	58	16	31	16	698	134	151	475	-3	
2	500	99.9	91	125	97	0	150	132	345	62	17	33	16	785	264	158	471	-41	
2	1000	99.98	15	251	107	0	300	132	345	63	18	34	16	947	2124	188	328	-191	

*These cases are referred back to the base case, but are predicted to achieve less than base case SO₂ removal. The \$/ton value listed represents a cost savings per ton of SO₂ not removed, compared to the base case.