JEWETT MINE DEWATERING: AN OVERVIEW

MARC J. NORRIS
SENIOR GEOLOGIST
NORTHWESTERN RESOURCES CO.
P. O. BOX 915
JEWETT, TEXAS 75848

ABSTRACT: The Jewett Mine is located near the town of Jewett in east central Texas. Mining depths range from 20' to 250'. Lignite produced at the Jewett Mine is from the upper portion of the Calvert Bluff Formation of the Wilcox Group.

For studies conducted prior to mining, dewatering was evaluated in terms of slope stability. Dewatering would be utilized to prevent massive circular slip failures that could impact the dragline operation which is the primary overburden removal system. The role of dewatering has been expanded to control piping, material flows, and in general, the amount of water in the pit.

Dewatering at the Jewett Mine started with 30 wells and has expanded to approximately 150 wells in 2 areas. Results of the dewatering efforts are evident in the pit, scraper operations, and water levels in observation wells.

Early studies provided a well field configuration for a given length of time. Logistical factors have affected the spacing. For a dewatering system to operate efficiently, many factors must be dealt with on a daily basis.

INTRODUCTION

The Jewett Mine supplies lignite to the Limestone Electric Generation Station of HL&P. Both the mine and power plant are located near the town of Jewett in east central Texas (figure 1). Mining depths range from 20' to 250'. Overburden is removed at the Jewett Mine by 3 Marion 8200 walking draglines and a fleet of 19 Caterpillar 637 scrapers. The dragline bench heights range from 20' to 103' depending on the depth of overburden, mode of operation and geotechnical considerations. Scrapers move the overburden not handled by the draglines. The life of mine area which is approximately 14,000 acres is divided into what is known as the West Half and East Half by a sand channel complex. Mining activities are presently confined to the West Half and specifically Areas A and C (figure 2). Shaded areas on figure 2 indicates dewatering activities. In full production, the Jewett Mine will deliver 8 million tons per year to the power plants. As mining progresses, so will the dewatering activities.

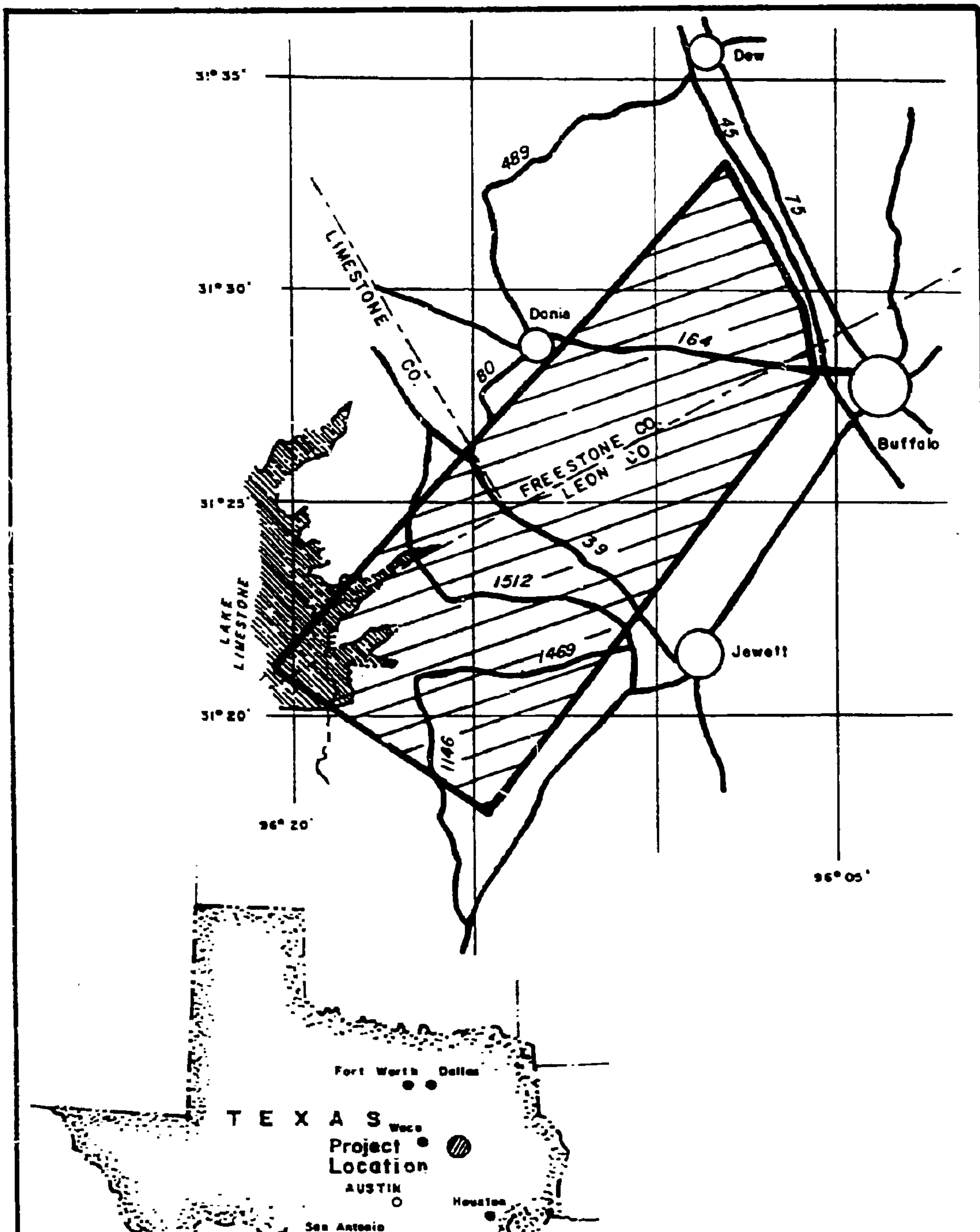

FIGURE 1

JEWETT MINE

GENERAL PROJECT LOCATION

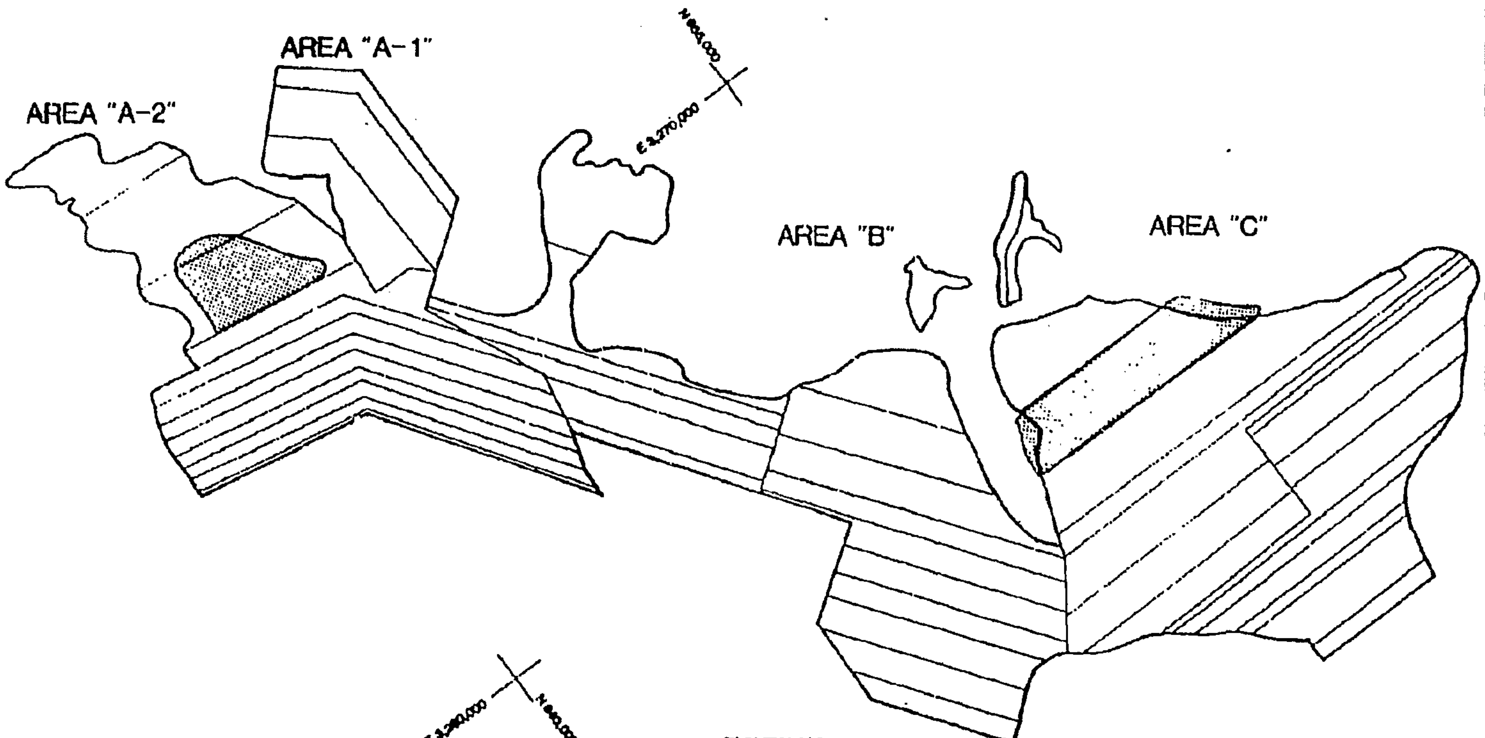

FIGURE NO. 2

JEWETT MINE-WEST HALF

STUDIES

State and Bureau of Economic Geology publications address the regional geology and lignite industry of Texas.

Studies conducted specificaliy for the Jewett Mine that pertain to groundwater include baseline as well as engineering studies. Baseline studies provided a precursory understanding of the aquifer characteristics of the Jewett Mine area. Engineering studies have included identifying potential mine hazards, geotechnical/hydrologic investigations and mine planning. The geotechnical/hydrologic studies related stratigraphy and water levels to highwall and spoil stability. Slope stability analyses provided one justification for dewatering but did not cover all aspects of dewatering. Dewatering studies for mine planning have been expanded to included controlling pit water as well as highwall failures. The basic well spacing was developed by means of a groundwater flow model.

STRATIGRAPHY

Lignite production at the Jewett Mine is from the upper portion of the Calvert Bluff Formation of the Wilcox Group. In the vicinity of the mine, the Calvert Bluff Formation is composed of sand, silt, clay and lignite with local ironstone concretions and channel lag gravels. In portions of the mine, the Carrizo Sand unconformably overlies the Calvert Bluff Formation. Both the Calvert Bluff Formation and Carrizo Sand are Eocene in age. The stratigraphy varies considerably from area to area at the Jewett Mine. The recovery schemes range from a single seam application to recovering as many as 4 seams from a single pit. Figure 3 is a generalized stratigraphic section showing the relative positions of the lignite seams.

Area A is a single seam operation. The dewatering target in Area A is a watertable sand that reaches over 100' in thickness. Area C is a multiple seam area. A watertable sand above the 3 seam and a confined sand between the 4 and 6 seams are the main targets in Area C.

MINING CONDITIONS

For the start up of the Jewett Mine, an attempt was made to handle groundwater in the pits. In Area A the ground water has been successfully controlled in the pit

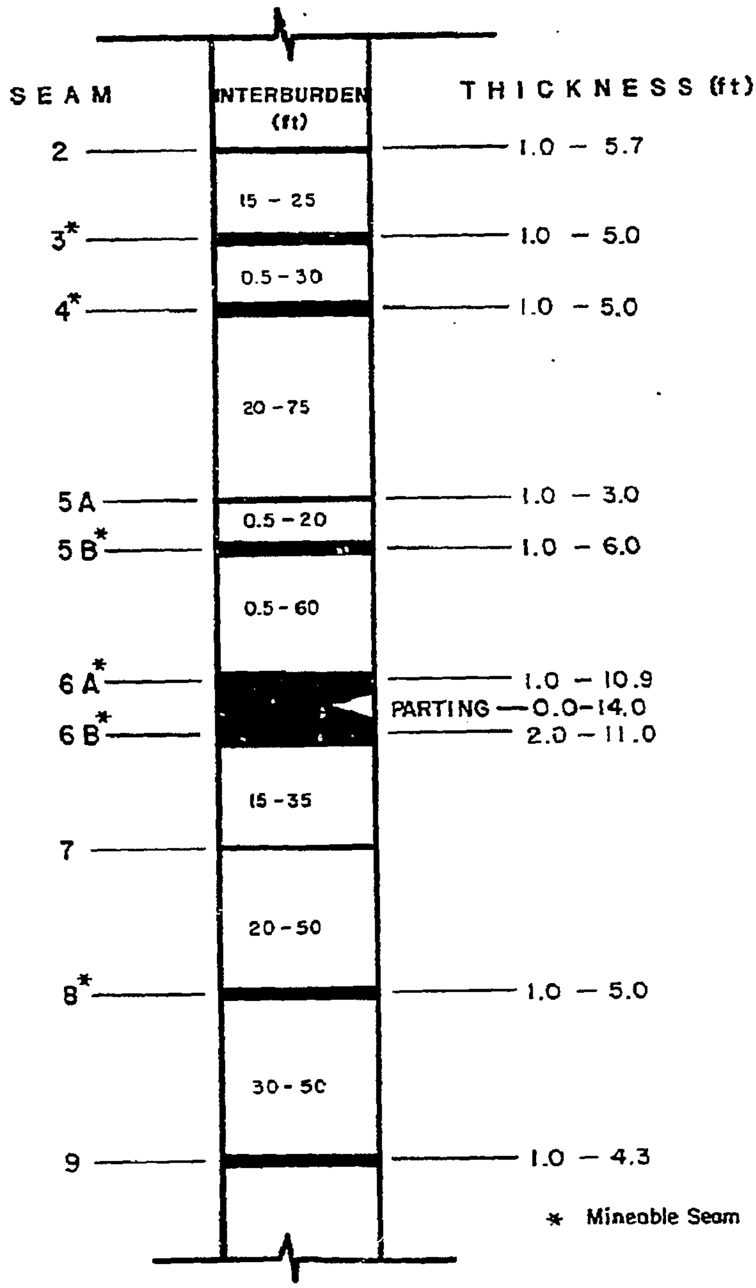

FIGURE NO. 3

General Stratigraphic Column

due to shallow mining depths and relatively low water levels. Portions of Area A-2 have thick saturated sand sequences requiring dewatering. Area A-1 was depleted and A-2 opened early in 1987.

The flow of groundwater into the Area C pit proved to be more than could be effectively handled in this manner. Thick sections of saturated sand in the highwall resulted in circular slip type failures. Piping, which is the rapid flow of groundwater into the pit, caused large cavities undercutting the highwall leading to sloughing type of failures. Saturated material placed in the spoil caused spoil side slides.

Coal recovery was hampered by water and mud in the pit. Slide material had to be rehandled to recover the coal. The slides created unacceptable pit conditions and an unsafe environment.

DEWATERING OPERATIONS

Massive dewatering efforts at the mine require a regularly spaced pattern of wells. The logistics of installing and servicing the wells and providing power and a pipeline discharge system lend themselves to a regular pattern of wells. The well field must blend with features of the mine such as service and haul roads, dragline power feed and fueling operations. As the dirt moving operations approach the dewatering system, the wells are retired one row at a time. Retiring an entire row of wells at one time saves rerouting of power feeds and discharge pipelines.

The initial installation was in Area C and consisted of 2 rows of 15 wells each. Wells were placed on 100' centers along the rows and the rows were spaced 120' apart. This type of operation was a "quick fix" for the Area C pit. Extended mine dewatering operations require lengthy pumping periods, and must be well in advance of the pit. The well field in Area C has advanced from the initial 30 wells to over 100 wells active at one time ahead of the prebench operation. Well installation began in Area A early in 1987. The well spacing is presently the same in both areas: 300' centers along the rows and 240' between rows. The rows of wells are staggered such that a well in one row bisects the distance between 2 wells in the next row as in figure 4. This type of spacing as opposed to a square grid reduces the diagonal distance between wells and creates more overlap of the cones of depression.

Observation wells are installed in each sand unit and sealed against leakage from other sands. Data generated from the observation wells yield information

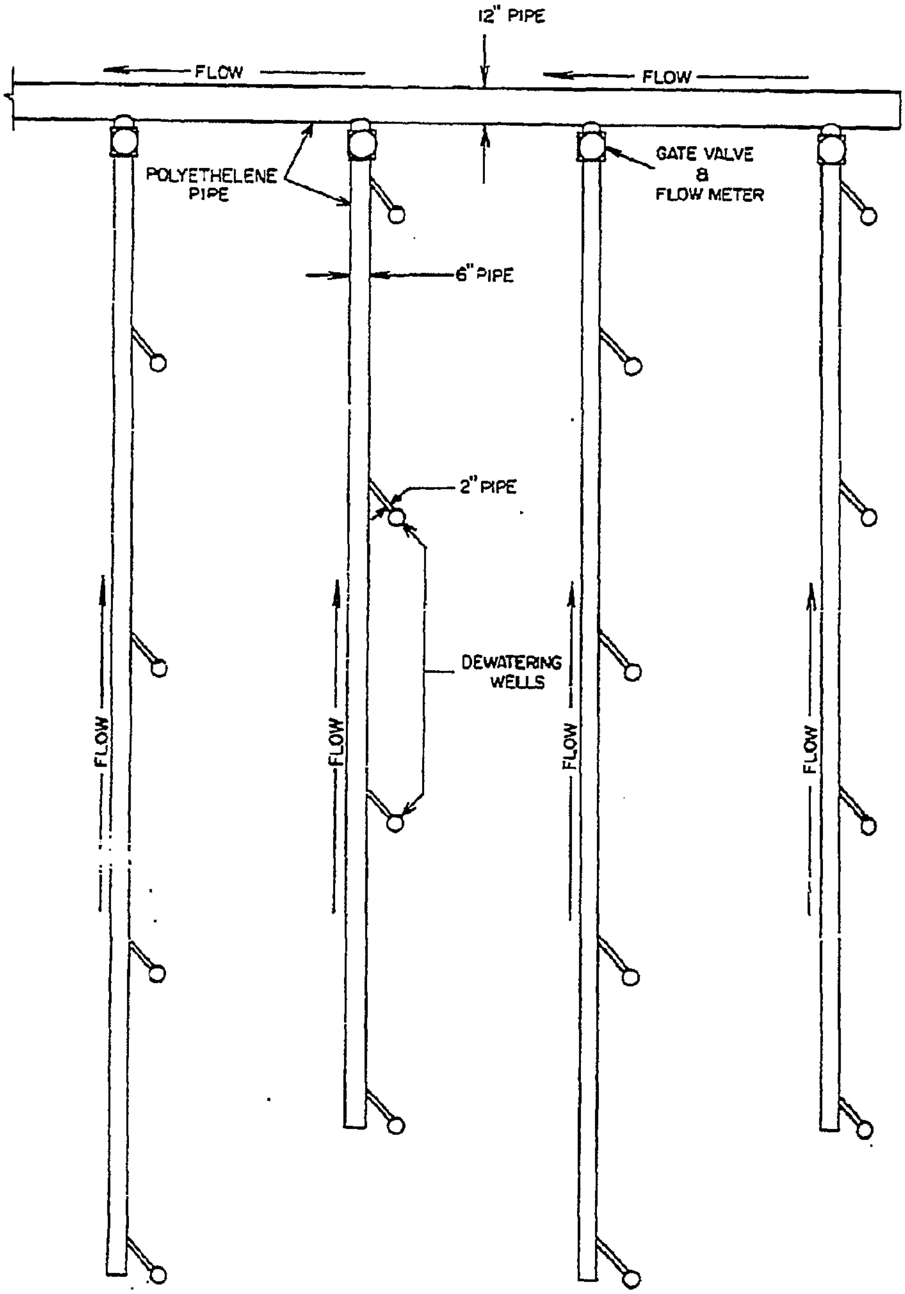

FIGURE NO. 4

TYPICAL WELL FIELD

concerning the effectiveness of the dewatering efforts on different levels. Water levels in discrete sands can have impacts on different operations such as prebench versus the dragline operation.

WELL INSTALLATION PROCEDURES

Dewatering wells are installed using both truck and buggy mounted mud rotary drilling rigs. Dug pits are utilized to provide as clean a hole as possible and drilling mud is seldom used. At each site, a 5 1/4" pilot hole is drilled and geophysically logged. The mine site hydrogeologist designs the well for each site from the drill cuttings and geophysical log. The pilot hole is stage reamed and the casing/slotted casing string is installed. Following casing installation the well is gravel packed and surged by the driller with a plunger on the bottom of the kelly. Bentonite is placed on top of the gravel pack and a surface seal is then set. Well development follows installation and consists of jetting the well with air. See figure 5 for a typical well.

WELL MATERIALS

Well casing consists of 6" schedule 40 PVC. Well screen is .016 mill slotted 6" schedule 40 PVC. The gravel pack is 16/30 silica sand. Centralizers are homemade items supplied by the drilling contractor. Drop pipe is threaded schedule 80, 1 1/2" PVC. Steel couplings are used for the PVC drop pipe. Figure 6 is an enlargement of the well head configuration.

Contractors are used for most aspects of the dewatering work except electrical. MSHA certification in low and medium voltage is required for electricians so that work is done in-house. The standard dewatering pump at the Jewett Mine is a 1.5 hp 3 phase 460 v all stainless steel submersible pump. Power is fed from the dragline substation to transformers which step the voltage from 22.9 KV down to 460V. Aluminum URD cable feeds power to the control panels at each well. In certain instances, the load demand and distance require a voltage boost installation.

In each panel is a contact which starts the submersible pump. A flow switch installed in the discharge line breaks the circuit when low flow is experienced. A timer energizes the starter contact after a prescribed amount of time.

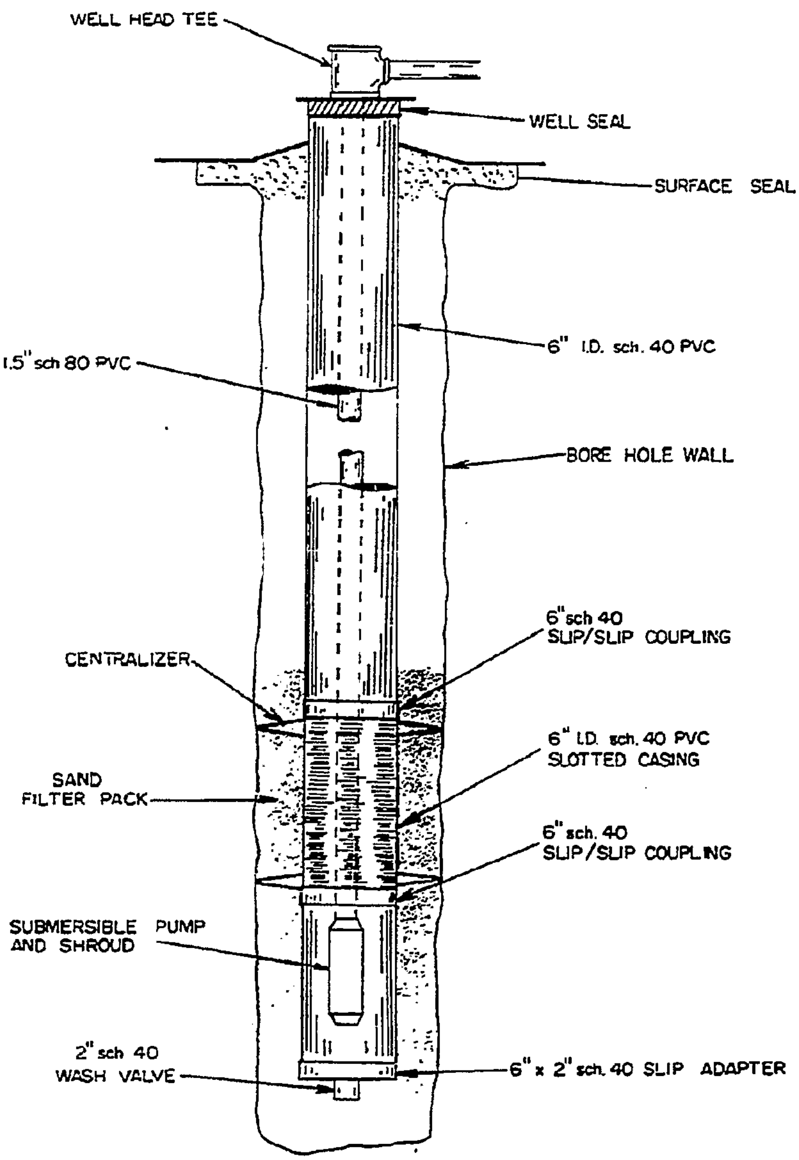

FIGURE NO. 5

TYPICAL DEWATERING WELL

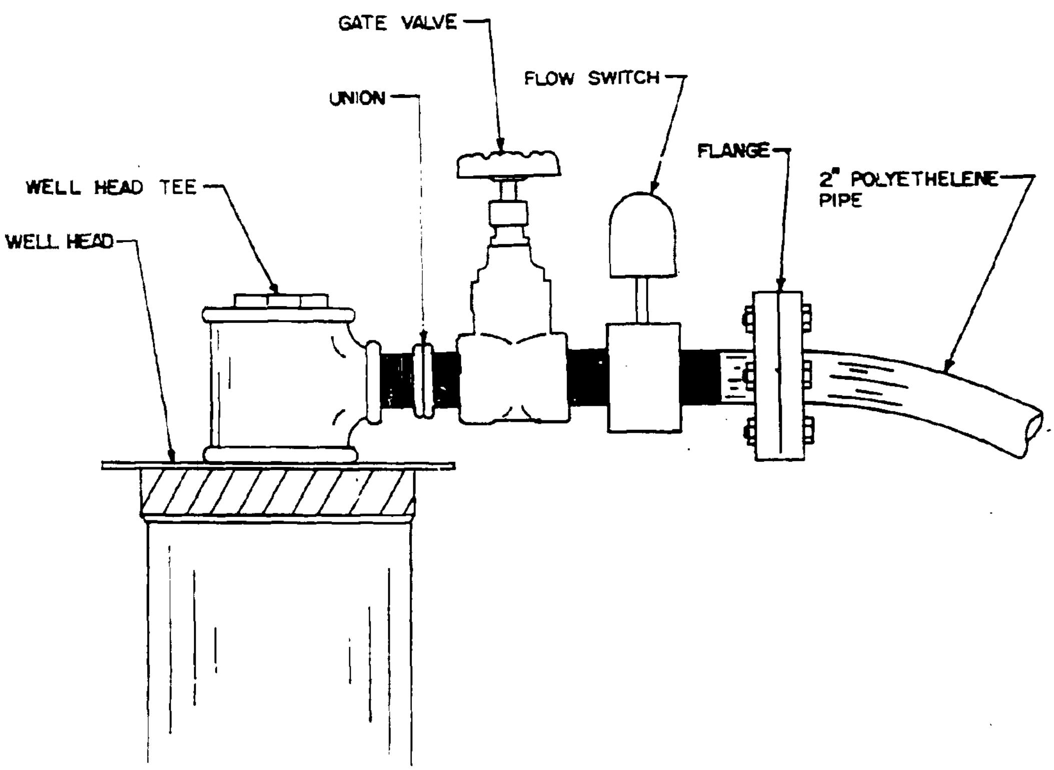

FIGURE NO. 6

DEWATERING HEAD EXTENSION DETAIL

The discharge pipeline is constructed of 2", 6" and 12" polyethelene pipe. Polyethelene pipe is joined by a butt fusion process and can be cut and rejoined at any point along the line. The polyethelene pipe can withstand light traffic. Breaks in the polyethelene pipe are often repaired by installing a full circle repair clamp. The polyethelene pipe is flexible enough to bend around obstacles and follow an irregular path when necessary. Fittings such as 45° or 90° bends, tee's, reducers and a variety of couplings are available for polyethelene pipe.

REAL WORLD PROBLEMS

An obstacle to the development of an efficient dewatering system, is tracking the performance of the well field and individual wells. Observation wells provide valuable information but do not document the performance of individual wells. Figure 4 shows flow meters which are planned for installation at the end of each row of wells. Regular monitoring of the meters will provide good estimates for total flow from the dewatering system. Hour meters to be installed in the panels will provide a record of how long each pump has operated since the last reading. The hour meters at the wells and flow meters at the end of each row of wells is not the optimum configuration for obtaining data, but is considerably less expensive than flow meters at each well. The total flow from a line of wells can be divided among the wells based on pump settings and hours pumped. This information is useful for planning well and pump maintenance schedules and identifying electrical problems. Drawdown verses time and flow information are vital to studies designed to optimize the dewatering system and extend dewatering activities to new areas. This information can be used to calibrate groundwater flow models and match well spacings to the time allotted for dewatering.

In some instances, installing wells on the designed pattern can be difficult, if not impossible. The well field plan may include ponds, pipelines, roads etc. Interference from topography and other features can be minimized if the well field is plotted on a topography base map, but field adjustments are still necessary.

Well and pump maintenance is a problem at the Jewett Mine. Iron in the groundwater precipitates as an oxide in the well, pump and drop pipe. Regular cleaning of the wells consists of brushing the screens and acid treatment. The screens are brushed with wire chimney brushes lowered and raised through the screened section by the pump contractor. The pumps and drop pipe are changed when the wells are serviced. Before the pumps are again deployed, they are soaked in an acidic solution. Steps have been taken to provide a well construction and pump placement that will prevent oxygen from reaching the pump and screens and inhibit the precipitation of the iron oxide.

Damage to the electrical distribution lines or pipeline discharge header can cause a shut down of the entire system. Gate valves have been installed as in figure 4 to provide the capability to isolate sections of the pipeline for repairs. Figure 6 is an expanded view of the well head. A gate valve and union are installed to allow the discharge line to be disconnected from a well without draining water onto the ground around the well.

A real challenge to the designer is to layout the system such that the opportunity for damage from prebench or topsoil removal operations is minimized.

The dewatering system has to respond to changes in mine plans. An insignificant change in plans for operations may be a major change for a dewatering plan. The person responsible for dewatering must keep abreast of operations and anticipate changes to the mine plan. Wells and electrical feed can be installed several months in advance of the anticipated service date to help reduce response time for dewatering. Stocking extra materials and pumps can reduce installation and repair time.

SUMMARY

Dewatering at the Jewett Mine involves a large number of closely spaced wells. The spacing was developed using a groundwater flow model. After reaching a conclusion on the well spacing, many decisions were made regarding materials and installation procedures. As the dewatering system expands, tracking performance of the wells becomes more involved. The dewatering system must be checked regularly for damage, failures and decreased well performance due to encrustation. Methods to improve the efficiency or decrease costs are constantly sought.

Enhancing Utility Baghouse Performance for Texas Lignite Ashes

by

Larry G. Felix, P. Vann Bush, and Randy L. Merritt

Southern Research Institute
P.O. Box 55305
Birmingham, AL 35255-5305

ABSTRACT

Utilities have come to expect clear stacks from the air pollution control devices associated with coal-fired boilers. Operating experience gained since the first installation of a utility baghouse in 1972 has shown that baghouses will provide this level of performance, even when alternative control device technologies cannot. Recently however, some utility baghouses have encountered particulate collection problems with the fly ash from certain Texas lignites from the Wilcox formation.

Detailed fly ash characterizations have indicated that the surfaces of the lignite fly ash particles are much smoother than fly ashes from low-sulfur western subbituminous or eastern bituminous coals, as well as being less cohesive. These features contribute to the formation of a filter cake with an inherently low porosity which tends to slowly bleed through typical bag fabrics. The result is a compound problem of high pressure drop and high stack opacity.

The Electric Power Research Institute (EPRI), as part of a program to develop and extend baghouse technology for utility applications, has sponsored research devoted to enhancing utility baghouse performance with this Texas lignite fly ash. This research has produced two methods to enable baghouses filtering this lignite fly ash to achieve particulate collection efficiencies that guarantee clear stacks. Neither approach leads to pressure drop penalties that could limit unit load. The first approach, flue gas conditioning with NH_3 gas, has provided this level of performance with conventional fiberglass fabrics. The principal effect of flue gas conditioning is to increase the cohesivity of the fly ash by the reaction of NH_3 gas with SO_3 present in the flue gas and on the fly ash particles. With the second approach, a conventional fiberglass fabric that has been coated on the filtering side with a microporous polytetrafluoroethylene (PTFE) film (Gore-Tex) has also provided this level of performance.

These two approaches have been tested in a year long fabric and flue gas conditioning study which was concluded recently at the TU Electric Monticello station Unit 1 baghouse. It has been determined that both approaches provide the capability of clear stack performance with no detrimental effect on baghouse operation, pressure drop, or on ash removal.

INTRODUCTION

Utilities have come to expect clear stacks from the air pollution control devices associated with coal-fired boilers. Operating experience gained since the first installation of a utility baghouse in 1972 has shown that baghouses will provide this level of performance, even when alternative control device technologies cannot (1). However, it has been established that baghouse performance depends on the filtration properties of the ash being collected (1-4). One case where baghouse performance has been clearly related to filterability of fly ash is at TU Electric's Monticello steam electric station, located near Mt. Pleasant Texas.

In this paper we review research carried out at the Monticello station that determined the cause for performance problems at the Units 1 and 2 baghouses. In addition, we describe approaches developed to significantly improve performance of these baghouses and present the results of compartment-level testing of each approach. One of these approaches, flue gas conditioning, represents the first application of a familiar technology for performance enhancement of electrostatic precipitators (ESPs) to a baghouse (2).

BACKGROUND

The Monticello station has three units fired with a locally strip-mined (Titus and Hopkins counties) Texas lignite coal. Units 1 and 2 are identical. Each is a 575-MW Combustion Engineering boiler followed by a 36 compartment Wheelabrator-Frye shake/deflate (S/D) cleaned baghouse that is paralleled with four small Research Cottrell ESPs. Unit 3 is a 750 MW boiler followed by an ESP and a wet scrubber. This paper deals only with Units 1 and 2.

When Units 1 and 2 came on line in 1974-75, they were equipped with four Research Cottrell ESPs per unit. Each boiler exhausted through two air heaters, and two ESPs were installed after each air heater. After startup both units exhibited excessive emissions. Flue gas conditioning did not sufficiently reduce emissions, and each unit was retrofitted with a 36 compartment S/D baghouse. The baghouses were originally sized to take 80% of the boiler exhaust at a net air-to-cloth (A/C) ratio of 2.9 acfm/ft^2 with an outlet mass concentration of no more than 0.009 grains/acf (0.024 lb/10^6 Btu). In this configuration each ESP would take about 5% of the total boiler exhaust.

Since coming on line in 1978-80, the Monticello baghouses have not performed like other utility baghouses associated with boilers fired with low-sulfur coals. These baghouses typically have a clear stack, flange-to-flange pressure drop that rarely exceeds 8 in. H_2O, and bag lives of four years or longer (1). The Monticello baghouses, however, regularly have had after-cleaning opacity spikes that exceed 20%, flange-to-flange pressure drops of 10-12 in. H_2O (figure 1), and bag lives of three years or less. Because of the high pressure drop, these baghouses cannot take more than about 50% of the boiler flow. This increases the amount of flow that each ESP must carry, which contributes a constant 5-10% constant baseline opacity.

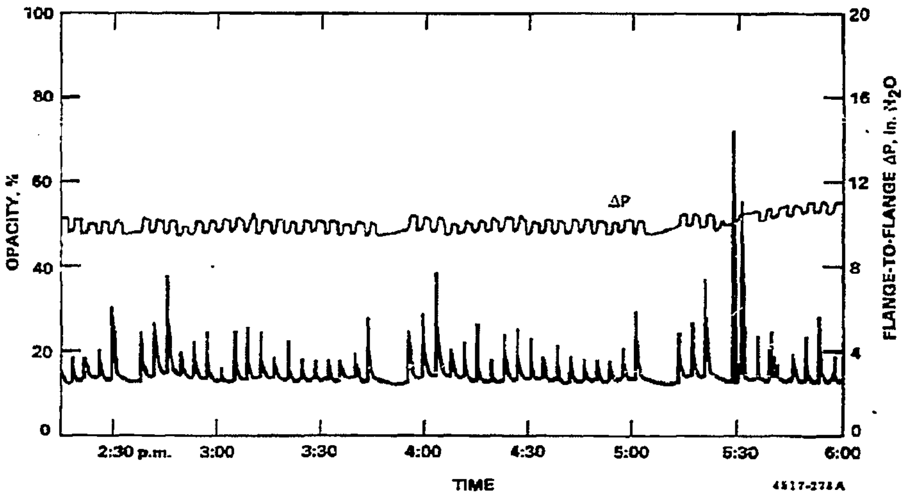

Figure 1. Flange-to-flange pressure drop and opacity for the Monticello Unit 2B baghouse, March 25, 1982, 2:30 p.m. - 6:00 p.m.; unit load was 58 MW.

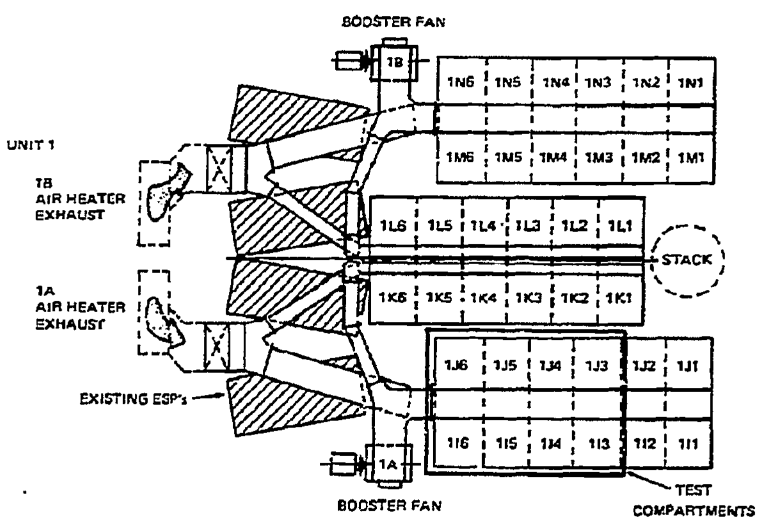

Figure 2. The Monticello station Unit 1 baghouse and ESP layout. Unit 2 has an identical layout.

Boilers

The 575 MW (net) Unit 1 and 2 Monticello CE boilers are supercritical, radiant, reheat steam generators with combined circulation. Each furnace is divided into two halves by a center wall with each half exhausting through an air heater. Table I shows the design fuel ultimate analysis (as received), the range expected and the results of an analysis carried out on a fuel sample from the Monticello Unit 2 coal supply. For comparison, results are also included from an analysis carried out on a fuel sample from the Public Service Company of Colorado's Arapahoe Station Unit 4 boiler which is the host boiler for the EPRI Arapahoe Test Facility and burns a low-sulfur, western, subbituminous coal. The Monticello lignite fuel has a very high ash and moisture content relative to that of the Arapahoe low-sulfur, subbituminous coal. This high ash content of the Monticello lignite results in very high mass loadings for flue gas exiting the Monticello boilers. Measurements at the air heater exit of the Unit 2 boiler gave an average value of 10.2 grains/acf.

Baghouses

The Monticello baghouses were provided by Wheelabrator-Frye, Inc. Each baghouse has 36 compartments with 204 bags per compartment and are S/D cleaned (figure 2). Table II gives detailed design specifications for the baghouses and the current bags. Figure 2 shows that each baghouse is made up of three 12 compartment baghouses and that the middle baghouse for each unit actually functions as two separate six-compartment baghouses. Thus the A and B air heaters each exhaust into 18 compartments of bags. Each set of 18 compartments is operated and cleaned independently. Presently a continuous cleaning cycle is used, with 18 compartments cleaned in 75 min. To minimize the opacity spikes associated with cleaning, the bags are shaken for only about 1 to 1.5 s.

Several types of filter fabrics have been tried at the Monticello baghouses (2). Currently, both baghouses are bagged with Menardi-Criswell 601-1-T fabric, with the warp side in (warp-in). This is a 9.5 oz/yd^2 fiberglass fabric woven in a 3 x 1 twill, with a 10% Teflon B finish. In such a 3 x 1 twill weave, 25% of the texturized fill yarns lie on the warp-in side of the fabric, and 75% of the texturized fill yarns lie on the other side of the fabric (the warp-out side). Thus this particular fabric has 25% of the side of the fabric which faces the dust covered with texturized or fuzzy yarns (25% exposed surface texturization) This fabric is widely used in utility baghouses, but usually in a warp-out configuration (75% exposed surface texturization). Bags made with warp side of the fabric facing the dust were chosen for the Monticello baghouses because they appeared to perform better than warp-out fabrics. Bag life with the warp-in fabric has been about three years. As with all of the fabrics which have been used at the Monticello baghouses, ash penetration into and through the fabric has been a continual problem which causes high pressure drops and large opacity spikes when a compartment comes on-line after cleaning.

Table I. As received fuel analysis for the Monticello and Arapahoe Stations.

	Weight, %			
	Monticello Unit 2			Arapahoe Unit 4
Component	Design	Typical Range	Measured (3/14/82)	Measured (1/13/82)
Moisture	31.8	26.4 – 36.5	30.04	11.93
Carbon	39.03	34.6 – 43.2	30.97	62.39
Hydrogen	2.98	2.6 – 3.4	2.56	5.02
Nitrogen	0.54	0.4 – 0.8	1.33	1.23
Chlorine	0.01	0.0 – 0.5	0.10	0.14
Sulfur	0.62	0.4 – 1.0	0.48	0.60
Ash	14.3	8.1 – 23.0	25.26	7.42
Oxygen	10.72	3.0 – 12.2	9.26	11.86
(Btu/lb)	(6711)	(5744 – 7883)	(5445)	(11015)

Table II. Design and operating data for the Monticello Units 1 and 2 Baghouses

Manufacturer	Wheelabrator-Frye Inc.
Cleaning Method	Shake/deflate
Design Parameters:	
Gas flow	1,840,000 acfm
Gas temperature	400°F
Inlet mass loading	10.0 gr/acf
Outlet mass loading	0.009 gr/acf
Flange-to-flange ΔP	7-8 in. H_2O
Air-to-cloth ratio, gross*	2.71 acfm/ft^2
Air-to-cloth ratio, net**	2.93 acfm/ft^2
Deflate gas air-to-cloth ratio	1.04 acfm/ft^2
Shake Frequency	4.0 Hz
Number of compartments	36, 18 per side
Number of bags per compartment	204
Total number of bags	7344
Bag diameter	11.5 in.
Bag length	30.75 ft
Collecting area:	
Per bag	32.6 ft^2
Per compartment	18,890 ft^2
Total	680,054 ft^2
Measured Parameters:	
Average flange-to-flange ΔP	11-12 in. H_2O
Average tube sheet ΔP	9-10 in. H_2O
Air-to-cloth ratio, net***	
10 in. H_2O tube sheet ΔP	2.30 acfm/ft^2
8 in. H_2O tube sheet ΔP	2.00 acfm/ft^2
6 in. H_2O tube sheet ΔP	1.59 acfm/ft^2
Drag (tube sheet ΔP/air-to-cloth ratio)	
10 in. H_2O tube sheet ΔP	4.35 in. H_2O·min/ft
8 in. H_2O tube sheet ΔP	4.00 in. H_2O·min/ft
6 in. H_2O tube sheet ΔP	3.77 in. H_2O·min/ft
Shake frequency	3.3 Hz
Shake amplitude	1.4 in., side-to-side
Shake duration	1-1.5 s
Bag cap acceleration	1.8 g, rms
Bags: (Now in Service)	
Manufacturer	Menardi-Southern
Fabric	601-1-T, warp-in
Finish	10% Teflon B
Weight	9.5 oz/yd^2
Months in service	36 maximum (Unit 1, K row)

*All compartments in service.
**One compartment per side cleaning (2 compartments), including deflation gas flow.
***575-MW unit load, Unit 2B baghouse, 18 compartments, 1 compartment cleaning.

RESEARCH

In 1982, the Electric Power Research Institute (EPRI) through its contractor for utility baghouse research, Southern Research Institute (SoRI), undertook a study of Monticello baghouses to determine the cause for their unusual behavior. On-site testing began in early in that year. Extensive testing was performed to characterize the behavior of the Monticello baghouses, coal, ash and dustcake samples were collected for laboratory analyses, and a small, portable Fabric Filter Sidestream System (FFSS) was used to study the filtration properties of the Monticello fly ash (2,3).

As this testing has been described in detail elsewhere, only the results will be reviewed here.

Results from Baghouse Tests:

1. The Monticello baghouses operate at an abnormally high flange-to-flange pressure drop. The opacity spikes associated with compartments returning to service after cleaning are not due to bag leaks or bag failures (figure 1).

2. Although the bag weights are quite low (0.2 to 0.3 lb/ft^2), samples of fabric removed from bags in the Monticello baghouses exhibit an unusually high resistance to air flow in either the forward (filtering) or reverse (cleaning) direction.

3. The interior of compartments at the Monticello baghouses are extremely dirty (2 to 8 in. deep in ash), and the outside of a typical bag is heavily coated with ash (about 1/8 in. thick).

4. Particle size distribution measurements made at the inlet of the Monticello Unit 2 baghouse showed that the Monticello ash is typical of that found at the exit of a pulverized coal-fired boiler.

Results from Laboratory Tests:

1. The Monticello fly ash has a chemical composition dissimilar to other low-sulfur coal ashes (1). In particular, it has a much higher calcium content than the fly ash from typical bituminous or subbituminous coal.

2. The Monticello fly ash has a low cohesivity compared to fly ash from non-lignite coals (2,5,6).

3. Scanning electron microscopy (SEM) conducted on Monticello fly ash revealed very smooth particle surfaces which are more free of submicron sized fly ash agglomerates than are typical for fly ash from non-lignite coals (1,2).

4. SEM examination of bag/dustcake samples that were preserved by encapsulation with epoxy indicated that the Monticello ash had so completely filled the fabric interstices that few flow channels remained (1).

Results from short-term (1 week) and long-term (1 month or longer) FFSS tests:

1. Typical Teflon B finished fiberglass fabrics were found to exhibit substantial amounts of ash bleedthrough, unlike ash from a bituminous or subbituminous coal (3).

2. Increasing the exposed surface texturization of fiberglass filtration fabrics tended to reduce ash penetration.

3. Gore-Tex (TM) membrane laminate on a standard fiberglass filtration fabric essentially eliminated ash penetration.

4. NH_3 flue gas conditioning at levels of 10 to 15 ppm greatly reduced ash penetration for all fabrics.

NH_3 flue gas was considered because the Monticello Unit 1 and 2 ESPs are conditioned with NH_3 (10-15 ppm) which reduces their emissions. However the SO_3 emissions are too low (1-2 ppm) to reduce the high ash resistivity (2×10^{12} ohm-cm at 360°F); instead, the NH_3 increases the cohesivity of the fly ash which reduces rapping emissions (7). Flue gas conditioning with NH_3 was evaluated because we suspected that it might increase the cohesivity of the Monticello fly ash to the point where a stable dustcake could be formed. To our knowledge, this work is the first case where it was shown that flue gas conditioning could substantially improve the performance of a fabric filter (2).

The above results led to the following general conclusions:

1. The atypical behavior of the Monticello baghouses is caused by the Texas lignite <u>fuel</u> and is not related to the design of the baghouses or boilers.

2. Flue gas conditioning with NH_3, Gore-Tex(TM) membrane laminates, and heavily texturized fabrics all appeared to be possible approaches to a solution of the Monticello baghouse problem.

LONG-TERM TESTS

At this point, a long-term (1 year) compartment-level testing program was recommended to evaluate each of the possible approaches listed above. After a project review meeting with TU Electric personnel in Dallas, it was decided to proceed with such a program at the Monticello Unit 1A baghouse. Four fabrics were selected for testing, each with and without NH_3 injection. These four fabrics included a heavily texturized fabric (approximately 100% of the exposed surface texturized), the fabric regularly used at the Monticello baghouses (approximately 25% of the exposed surface texturized - a warp-in 3 x 1 twill fabric), the latter fabric turned inside-out (approximately 75% of exposed surface texturized - the fabric most commonly used in utility baghouses), and a Gore-Tex (TM) laminate material on a fiberglass filter fabric. Table III describes each of these fabrics in more detail. Figure 3 shows SEM photographs of each of these fabrics.

Table III. Fabrics Selected for Long-Term Testing.

Fabric[1]	Description	Compartment	Date Installed
MC 601-1-T (warp in)	3 x 1 twill weave, 54 x 30 count, 150-1/2 warp, 150-1/4 TEX fill, 9.5 oz/yd^2, 10% Teflon B finish. 25% exposed surface texturization.	1I4 1J3 1J4	6/4/85 6/7/85 6/6/85
MC 601-1-T (warp out)	3 x 1 twill weave, 54 x 30 count, 150-1/2 warp, 150-1/4 TEX fill, 9.5 oz/yd^2, 10% Teflon B finish. 75% exposed surface texturization.	1I3	6/4/85
MC 996T (warp out)	Double warp weave, 50 x 30 count, 37-1/0 TEX + 37-1/0 warp, 75-1/3 TEX fill, 16 oz/yd^2, 10% Teflon B finish Near 100% exposed surface texturization.	1I5 1J5	5/31/85 6/5/85
Gore-Tex[2]	3 x 1 twill weave, 9.8 oz/yd^2, Gore-Tex laminate, 10% Teflon B finish.	1I6 1J6	5/30/85 9/9/85 5/30/85 4/25/86

[1]MC = Menardi Criswell Corp.
[2]Both 1I6 and 1J6 were rebagged after stitching problems were discovered in the Gore-Tex bags.

Table IV. Results from compartment-scale tests at the Monticello Unit 1A baghouse

	Test Data by Compartment							
	1I4	1J3	1J4	1I3	1I5	1J5	1I6	1J6
Fabric Type:	601-1-T	601-1-T	601-1-T	601-1-T	996T	996T	Gore-Tex®	Gore-Tex®
Texturization (%):	25	25	25	75	100	100	0	0
Date on Line:	6/4/85	6/7/85	6/6/85	6/4/85	5/31/85	6/5/85	5/30/85*	5/30/85**
NH_3 Injection:	No	Yes	Yes	No	No	Yes	No	Yes
	Average Tube Sheet ΔP,*** in. H_2O							
9/85	9.1	8.9	8.9	9.2	9.0	8.3	7.7	7.4
5/86	8.8	8.6	8.6	8.9	7.7	8.5	7.2	7.6
	Average Drag Relative to Compartment 1I4							
9/85	1.0	0.81	0.78	0.92	0.91	0.67	0.56	0.51
5/86	1.0	-	0.76	1.07	0.86	0.87	0.51	0.53
	Mass Emissions, 9/85							
gr/acf x 10^3	3.88	0.60	0.89	4.66	1.53	0.21	0.58	0.26
(lb/10^6 Btu) x 10^3	10.31	1.60	2.36	12.4	4.08	0.55	1.53	0.70
	Residual Dustcake, (lb/ft^2)							
9/85	0.16	0.17	0.28	0.25	0.23	0.20	0.07	-
1/86	0.33	0.44	0.43	0.48	0.30	0.40	0.27	0.08
5/86	0.27	0.39	0.55	0.43	0.32	0.38	-	0.10

*Rebagged on 4/25/86
**Rebagged on 9/5/85
***Referenced to a flange-to-flange ΔP of 10 in. H_2O.

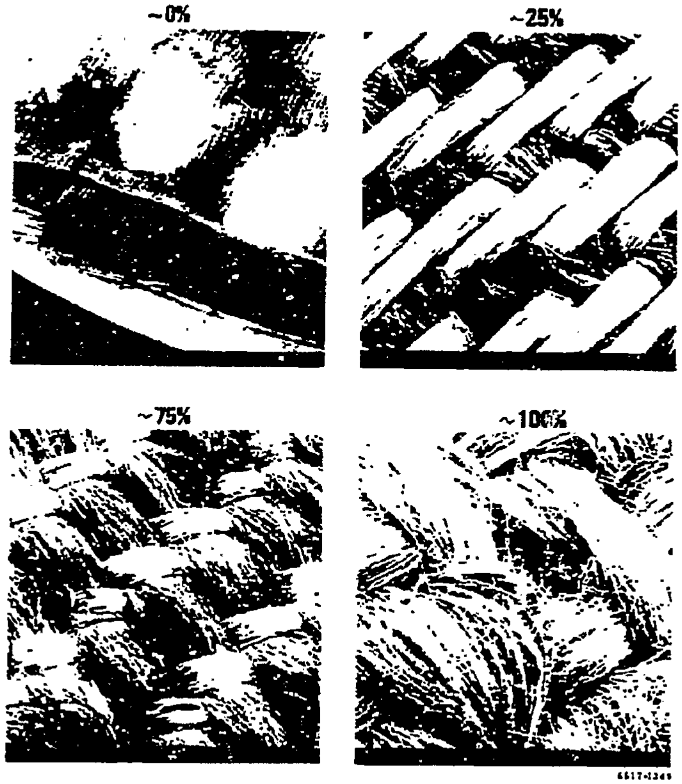

Figure 3. Fabrics of differing levels of texturization tested at Monticello.

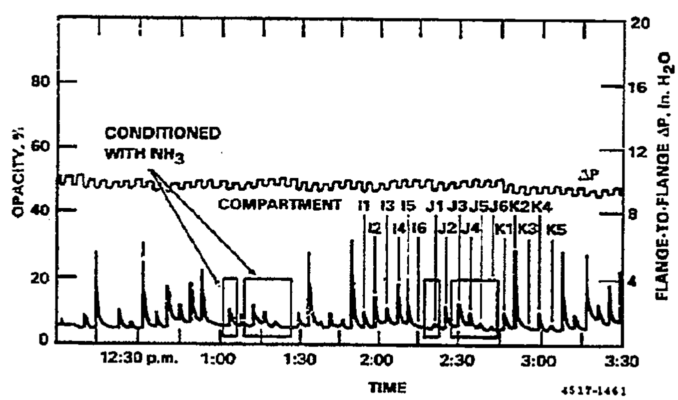

Figure 4. Flange-to-flange pressure drop and opacity for the Monticello 1A baghouse, May 15, 1986, 12:30 p.m. - 3:30 p.m.; unit load was 600 MW.

Four compartments in the I row (1I3 - 1I6) and four compartments in the J row (1J3 - 1J6) of the Unit 1A baghouse were rebagged with the test fabrics. NH_3 injection systems were fabricated and installed on the inlet elbows of the four compartments on the J row, and the test fabrics were arranged so that compartments that faced each other across the two rows were bagged with the same fabric. Table III lists the compartments in which the fabrics were installed and gives the installation date for each fabric. The NH_3 injection systems were calibrated for a nominal delivery rate of 15 ppm. Through an error, compartment 1J3 was rebagged with the 25% rather than 75% exposed surface texturization fabric. This error was not discovered until the test program had begun.

When the test compartments came on line, and at various periods since that time, the performance of the test compartments was measured by monitoring particulate concentrations inside the compartments and by recording tube sheet and flange-to-flange pressure drop. After several months, particulate emissions in all of the test compartments had stabilized at low levels, so later performance measurements involved the measurement of relative compartment drag (drag relative to compartment 1I4 which did not have NH_3 injection and was bagged with the fabric commonly used at the Monticello baghouses) and dust cake weights (areal density of the permanent dustcake). Drag is defined as the ratio of tubesheet pressure drop (pressure drop across the bags in a compartment) to air-to-cloth ratio (the volumetric flow of gas through a compartment of bags divided by the effective filtering area of the bags in the compartment). These measurements are summarized in Table IV.

In the absence of flue gas conditioning, it is evident that ash slowly bleeds through the bags in the Monticello baghouse. When the bags are new, and the dust has not permeated the fabric, there are many pores in the fabric through which the non-cohesive dust tends to sieve. Later, when the dust has filled the fabric, mass emissions are much less. In NH_3 conditioned compartments, the emission rates are always low, but even these emission rates decrease as the bags season and stable dust cakes are formed. Measurements of mass emissions throughout a filtration cycle and average tubesheet pressure drop made in compartments 1I3 (no NH_3) and 1J3 (with NH_3) illustrate this behavior:

	Emissions ($lb/10^6$ Btu)		Tubesheet ΔP (in. H_2O)	
	Compartment			
Date	1I3	1J3	1I3	1J3
6/85	0.167	0.012	-	-
7/85	0.010	0.006	8.1	8.0
9/85	0.012	0.002	9.2	8.9

Figure 4 shows a flange-to-flange pressure drop and opacity record for the unit 1A baghouse from May 15, 1986 with compartments indicated which were being conditioned with NH_3. At this time compartment 1J1 was also being conditioned with NH_3 to study effect of NH_3 conditioning on bags which had been in service for long periods of time (three years) before encountering NH_3 for the first time. In compartment 1J1 the opacity spike was greatly reduced, but drag was not lowered.

Real-time measurements with an optical particle counter indicate that most of the mass emissions (approximately 80 - 90%) occur during the opacity spike. Considering the above measurements and Figure 4, the mass content in an opacity spike must be very low.

After more than one year of operation, the NH_3 injected compartments have, overall, performed better than the compartments that were not conditioned with NH_3. Bag weights have been higher for the ammonia injected compartments, which is due to non-optimum cleaning and increased cohesivity of the NH_3 conditioned ash. On-line measurements shown in Table IV and Figure 4, and off-line inspections have shown that flue gas conditioning with NH_3 reduced:

- Particulate emissions
- After-cleaning opacity spikes
- Drag, for standard bag fabrics, and
- Ash bleedthrough

The following general observations can be made (with or without NH_3 injection):

- The best performing fabric is the Gore-Tex membrane fabric.
- Higher degrees of surface texturization texturization appear to reduce emissions with little drag penalty.

Finally, recent laboratory measurements on fly ash samples from the interior of bags in compartments with and without NH_3 conditioning have shown that the bulk porosity and cohesivity of NH_3 conditioned Monticello ash are measurably greater than that of the unconditioned ash, as is shown in Table V. This table also shows that ash from bags that have been conditioned with NH_3 contain larger amounts of soluble $SO_4^=$ and much larger amounts of soluble NH_3 compounds than do unconditioned ash samples. The cohesivity of Monticello fly ash that was conditioned with NH_3 is equivalent to the cohesivity of ashes from baghouses associated with boilers fired with low-sulfur subbituminous coal (8).

CONCLUSIONS

High pressure drop and opacity spikes at the Monticello baghouses were found to be fuel-related and are not caused by the boiler or baghouse design. The fuel used at the Monticello station yields a fly ash with a very low cohesivity.

This low cohesivity allows the ash to impregnate the filter bags, which results in a high pressure drop. When a compartment comes on line after cleaning, ash which has bled through the dustcakes and filter bags creates large opacity spikes.

Research was directed toward improving the performance of the Monticello baghouses to equal the performance of other utility baghouses associated with boilers fired with low-sulfur coals. Two methods for improving performance have been found:

Table V. Results of Laboratory Measurements Made on Monticello Fly Ash

A. Cohesivity Related Measurements

Sample Description	Effective Angle of Internal Friction[1] (°)	Bulk Porosity,[2] Heated - (350°) (%)
With NH_3 Injection	41.5	42.0
Without NH_3 Injection	38.5	38.4

B. Soluble Compound Measurements

Sample Description	Soluble $SO_4^=$ (% weight)	Soluble NH_3 (% weight)
With NH_3 Injection	0.69	0.15
Without NH_3 Injection	0.48	0.00049

[1] The effective angle of internal friction is a measure of relative cohesivity. The higher the angle, the greater the cohesivity (5,6)

[2] Calculated assuming a particle density of 2.4 gm/cm^3.

- Flue gas conditioning with NH_3 gas was found to increase the cohesivity of the Monticello ash so that it could be removed from the flue gas with conventional woven filtration fabrics.

- Both heavily texturized, woven fiberglass filtration fabrics and woven fiberglass fabrics coated a Gore-Tex membrane were found to filter Monticello ash efficiently.

Both of these approaches to improving the performance of the Monticello baghouses have yielded promising results and both appear to be equally viable.

Flue gas conditioning is particularly interesting because it appears to be an effective means of transforming an essentially unfilterable ash into one that is easily filtered by most filter fabrics. Since there are other utility coal-fired boilers fueled with low-rank coals similar to that used at the Monticello station, those utilities could consider the installation of a fabric filter with the knowledge that it could meet any foreseeable particulate emissions standard.

ACKNOWLEDGMENTS

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FLUE GAS CONDITIONING FOR BAGHOUSE PERFORMANCE IMPROVEMENT WITH LOW-RANK COALS

D.L. Laudal and S.J. Miller

University of North Dakota Energy Research Center

ABSTRACT

Many low-rank coals produce high-resistivity fly ashes, which makes electrostatic precipitation unattractive for these fuels. As a result, fabric filtration has become a choice method of particulate control for low-rank coals. Fabric filtration performance, however, is dependent on the physical and chemical properties of the fly ash. Some low-rank coals have been identified, in both the pilot-scale and full-scale tests, to produce fly ash which is difficult to collect with conventional fabric filtration. The exact fly ash characteristics which cause collection problems in a baghouse are not well known, but are thought to include particle size and cohesiveness. Ashes which are more difficult to collect have a low cohesiveness compared to other ashes.

Pilot-scale tests have been conducted at the University of North Dakota Energy Research Center in which small amounts of ammonia and sulfur trioxide were injected into the flue gas upstream of the baghouse. Results show that, for several low-rank coals which produce difficult-to-collect ash, particulate emissions were substantially reduced when using these conditioning agents. As a bonus, baghouse pressure drop was also reduced making the process much more attractive. Since injection technology for conditioning agents such as ammonia and sulfur trioxide has been well developed in the electrostatic precipitator industry, this method is readily available. Longer term and larger scale testing, however, is needed to more fully evaluate the potential of conditioning agents to improve fabric filter performance.

An economic analysis shows that the use of ammonia and sulfur trioxide as conditioning agents would add about 9% to the cost of operating a baghouse. However, this cost could be more than recovered if the size of the baghouse could be reduced because of the lower expected pressure drop.

INTRODUCTION

Fabric filters are being applied to utility coal-fired boilers in increasing numbers, particularly in the West where high ash resistivity has made electrostatic precipitators (ESPs) less attractive. However, major questions exist as to the cause and effect relationships between ash properties and emission levels, as well as pressure drop. Although most baghouses are able to meet or exceed current New Source Performance Standards, high emissions from a utility baghouse have been reported (1). In addition, in recent years there has been increased concern over the effects of fine particle emissions on health.

Results from previous tests at the University of North Dakota Energy Research Center (UNDERC) in a pilot-scale combustor and baghouse have shown that baghouse particulate emissions are dependent upon the ash composition (2,3). This suggests that changing the ash composition with additives may improve baghouse collection efficiency.

Exploratory screening tests were conducted to evaluate ammonia and SO_3 as conditioning agents for reducing fine particulate emissions for difficult-to-collect ashes. Initial tests, with ammonia and SO_3 individually did not result in a significant improvement. However, when ammonia was used in combination with SO_3, there was a significant reduction in fine particulate emissions. Initial results of the effect of conditioning agents on baghouse performance were previously reported (4,5). This paper summarizes earlier results and presents additional data from research conducted in the past year.

EXPERIMENTAL

Tests were conducted with the UNDERC particulate test combustor (PTC) and pilot-scale baghouse. The pilot furnace is a 550,000 btu/hr pulverized coal-fired unit designed specifically to generate fly ash representative of that produced in a full-scale utility boiler. The pilot baghouse includes separate chambers, which allow operation in either a pulse jet or shaker cleaning mode. The pulse jet chamber bags are cage mounted while the shaker chamber has tube sheet mounted bags. The nine bag shaker chamber, which was used for these tests, has three compartments with three bags in each. The fabric used was a conventional 10 oz woven glass with 10% teflon B finish. The air-to-cloth ratio for all of the tests conducted using the shaker chamber was 3 to 1. During the earlier tests, three bags were taken off-line and cleaned every 20 minutes. For the recent tests all 9 bags were taken off-line and cleaned simultaneously every 2 hours. The baghouse temperature for all the tests was 300°F.

The method used to generate SO_3 was to catalytically convert SO_2 to SO_3 using vanadium pentoxide (V_2O_5). Air and SO_2 are passed through a V_2O_5 filled reactor that is heated to 800°F, oxidizing the SO_2 to SO_3. The percent conversion was measured using SO_2 analyzers and was found to be greater than 99%. Therefore, the quantity of SO_3 injected was controlled by using a flow meter to control the SO_2 to the generator. The ammonia was injected into the flue gas just upstream of the SO_3 injection point at a temperture of 700°F.

The test program was divided into two phases. Phase I consisted of initial tests on 3 low-rank coals (4). The tests were conducted primarily to determine if SO_3 and ammonia conditioning is an effective method for improving fine particulate collection efficiency in a baghouse and to determine what effect it may have on operating pressure drop. After the Phase I initial evaluation tests were completed, the following questions remained:

1. Will the process work at lower injection temperatures (temperatures equivalent to the cold side of an air preheater)?
2. Can a lower concentration of ammonia be used?
3. What is the effect of coal type (e.g. higher rank coals)?

Phase II tests were designed to help answer these questions.

Two of the coals chosen for the Phase I tests, Falkirk, North Dakota lignite and Big Brown, Texas lignite were selected because earlier tests (2,3) with these coals resulted in high fabric filter emissions. The third coal, South Hallsville, Texas lignite was chosen because its ash composition indicated that this coal would also produce a fly ash that would be difficult-to-collect in a fabric filter (2). For each coal a baseline test was conducted without conditioning. An identical test was performed with SO_3 and ammonia conditioning to determine differences in emissions and pressure drop. Tests with these coals were completed with the baghouse shaker chamber starting with new bags. The first 2-3 hours of each test were performed without bag cleaning to measure both particulate emissions and pressure drop as a function time with dust cake build-up on the fabric. Bag cleaning was then initiated, and over several cleaning cycles, emissions as a function of time as well as total dust loading measurements were taken. In addition, two 50 hour tests utilizing South Hallsville lignite were completed to evaluate the longer term effects of conditioning on emissions and pressure drop.

An injection location well upstream of the baghouse was chosen for Phase I testing to provide enough residence time for adequate mixing of the SO_3 and ammonia with the flue gas. At this injection point, the flue gas temperature was 700°F. Additive concentrations in the flue gas for these tests were 12 ppm SO_3 and 45 ppm ammonia. However, one short test was conducted where the concentrations were increased to 60 ppm SO_3 and 225 ppm ammonia to help evaluate the reaction products.

For Phase II tests, three different coals were chosen. The first coal, Monticello, Texas lignite, was selected because this coal produced an ash that was difficult to collect in a full-scale baghouse (1). The other two coals selected were Sarpy Creek, Montana subbituminous, and an Indiana bituminous coal. Three runs were completed with Monticello lignite including a baseline test, a conditioning test with high temperature injection (700°F), and a conditioning test with low temperature injection (350°F). Two tests were conducted with each of the other coals, a baseline test, and a low temperature injection test. For all of the tests, the ammonia and SO_3 injection rates were set at 25 and 12 ppm, respectively. To provide a better understanding of dust cake build-up and its relationship to penetration, the cleaning cycle for the Phase II tests was changed from that used during Phase I. Starting with new bags the first three hours were performed without cleaning, after which all of the bags were cleaned off-line every two hours.

A Flow Sensor, six-stage multicyclone was used to measure the inlet fly ash particle size distribution as well as to collect size fractionated samples for subsequent analyses. EPA method 5 was used to determine the inlet and outlet dust loadings for each test to calculate the overall baghouse efficiency. Particulate emissions from the baghouse were measured in real-time with a TSI APS 33 aerodynamic particle sizer. The APS 33 has the capability to measure particles in the size range of 0.5 to 15 µm and has been described in a previous report (6). Submicron particles (0.01 to 1.0 µm) were measured using a TSI Differential Mobility Particle Sizer (DMPS). This instrument separates various size particles using electrical mobility, which is dependent on particle size, and then counts the number of particles in the resultant monodispersed aerosol with a Condensation Nucleus Counter (CNC). Since the DMPS can take up to 25 minutes for a measurement, the CNC, with an impactor to remove particles greater than 1 µm, was used to obtain a real-time submicron particle count. The particulate sampling scheme is shown in Figure 1.

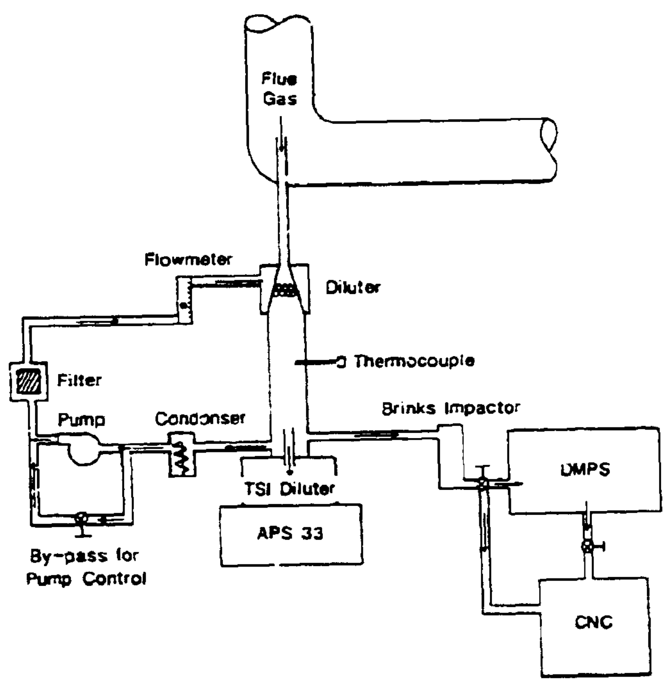

Figure 1. Schematic of particulate sampling system.

RESULTS AND DISCUSSION

Baghouse Emissions

Particulate removal efficiencies were determined by inlet and outlet dust loading measurements. The dust loadings were taken spanning several cleaning cycles using woven glass bags and shaker cleaning. The results are summarized in Table 1 for both Phase I and Phase II tests. A significant improvement in removal efficiency is clearly seen using SO_3 and ammonia conditioning. Removal efficiency for a Beulah lignite is included as a reference since, prior to flue gas conditioning, this coal resulted in the highest removal efficiency of all coals tested (3). The results from Phase I cannot be compared directly to those from Phase II, since the cleaning cycle was different. During Phase I, the outlet dust loadings included 3 to 4 20 minute cleaning intervals, while the dust loadings from Phase II were completed between bag cleanings. Therefore, the Phase I dust loadings included the increased emissions that occur immediately following bag cleaning. However, for both Phase I and II, the dust loading measurements show significant improvement in removal efficiency.

The APS 33 measures emissions as a function of time for particle sizes from 0.5 to 15 µm, but to avoid multiple graphs for different particle sizes, the respirable mass function of the APS was used. This is a weighted sum of the mass of particles between 0.5 and 10 µm, including all the mass of particles from 0.5 to 1.5 µm and a decreasing percentage of the mass of those between 1.5 and 10 µm. The calculation of mass concentration from number concentration data assumes spherical particles and requires an input of the fly ash particle density. Figure 2 shows the respirable mass emissions from the shaker chamber with no cleaning cycle, starting with new bags for the three lignites tested during Phase I. With conditioning, respirable mass emissions quickly drop for the first 40 minutes of dust cake build up and then appear to remain at a constant low level. Respirable mass emissions after 40 minutes are reduced by 3 to 4 orders of magnitude by conditioning with SO_3 and ammonia. The particulate removal efficiency for these three coals after two hours is greater than 99.999%.

Figure 3 illustrates the cleaning cycle effects for South Hallsville lignite with and without conditioning. As expected, just after shaking the bags, there is a significant increase in emissions for the conditioning tests but there is also a rapid recovery. Throughout several cleaning intervals there is a significant reduction in fine particle emissions. The same results have been reported for the Big Brown and Falkirk lignites (4). Figures 4 to 6 show the effects of bag cleaning on emissions using Monticello lignite, Indiana bituminous, and Sarpy Creek subbituminous. During Phase II tests, all of the bags were cleaned off-line every 2 hours following the initial 3 hour dust cake build-up. For the Monticello lignite and the Indiana bituminous coal, there was a very significant decrease in respirable mass emissions with conditioning. However, for Sarpy Creek coal the decrease was not as significant. This may be a result of differences in ash chemistry as will be discussed later in the report.

TABLE 1

PERCENT BAGHOUSE REMOVAL EFFICIENCY DETERMINED BY EPA METHOD 5 DUST LOADINGS

Coal	Nonconditioned	Conditioned
(Phase I - 20 min. cleaning interval)		
Beulah	99.93	--
Falkirk	94.70	99.64[a]
Big Brown	80.00	99.95
South Hallsville	91.03	99.90
South Hallsville[b]	99.03	99.76
(Phase II - 2 hour cleaning interval)		
Monticello	96.57	99.98[c]
Monticello	96.57	99.97[d]
Sarpy Creek	99.82	99.92
Indiana Bituminous	99.41	99.99

[a] A leak between compartments caused some forward flow through bags while shaking. Removal efficiency would likely be higher if the leak had not occurred.

[b] Pulse jet results with felted fabric.

[c] High temperature injection of conditioning agents (700°F).

[d] Low temperature injection of conditioning agents (350°F).

Using the CNC and DMPS, the submicron particles (0.01 to 1.0 µm) were monitored for the South Hallsville tests during Phase I and for all the tests during Phase II. The South Hallsville tests show that submicron particle emissions were also reduced at the baghouse outlet by over 3 orders of magnitude when comparing the conditioned to the nonconditioned results. These instruments were used more extensively for Phase II testing and again, SO_3 and ammonia conditioning resulted in a substantial reduction in submicron emissions for all coals with the exception of Sarpy Creek. For this coal, submicron particulate emissions were slightly less for the conditioned test compared to the baseline test. Flue gas conditioning has been shown to be an effective method of reducing fine particulate emissions from a baghouse for all lignites tested. For the two higher rank coals tested, the bituminous coal showed a substantial reduction in emissions with conditioning. However, the subbituminous coal tested did not show as great an improvement.

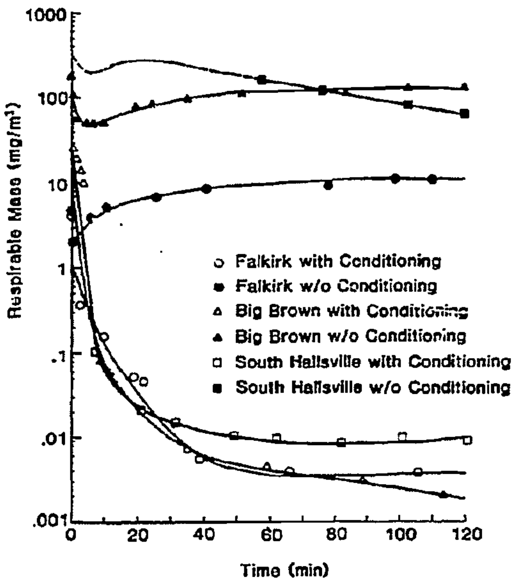

Figure 2. Phase I: Respirable mass emissions as a function of time without cleaning bags.

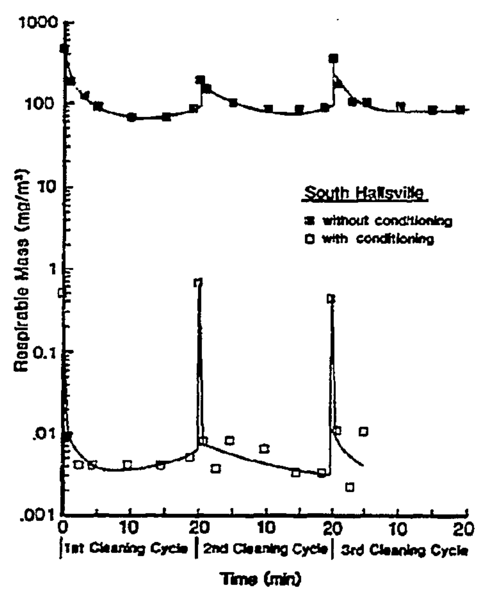

Figure 3. Respirable mass emissions (20 min shaking interval) with South Hallsville, Texas lignite.

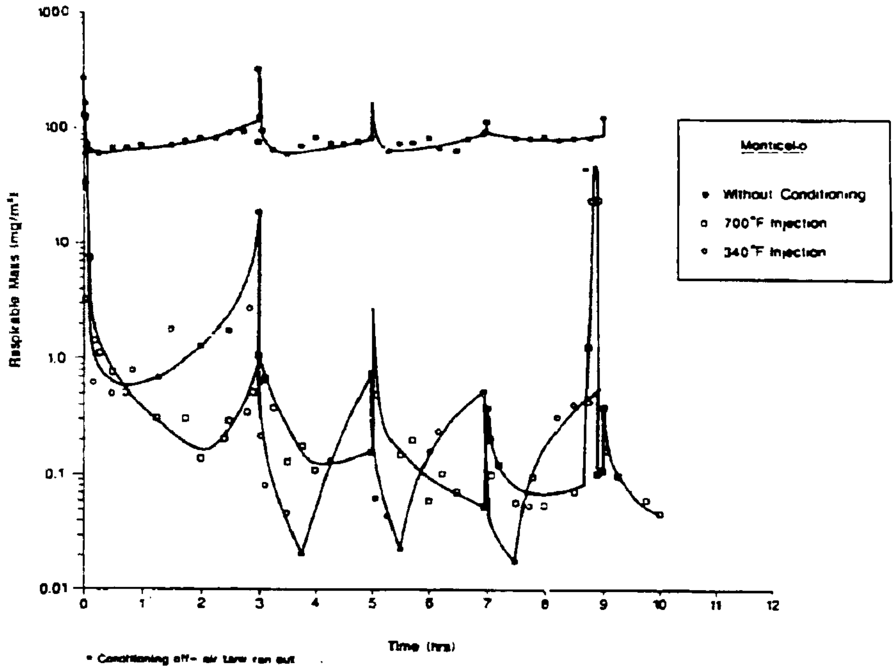

Figure 4. Respirable mass emissions (2 hour shaking interval) with Monticello, Texas lignite.

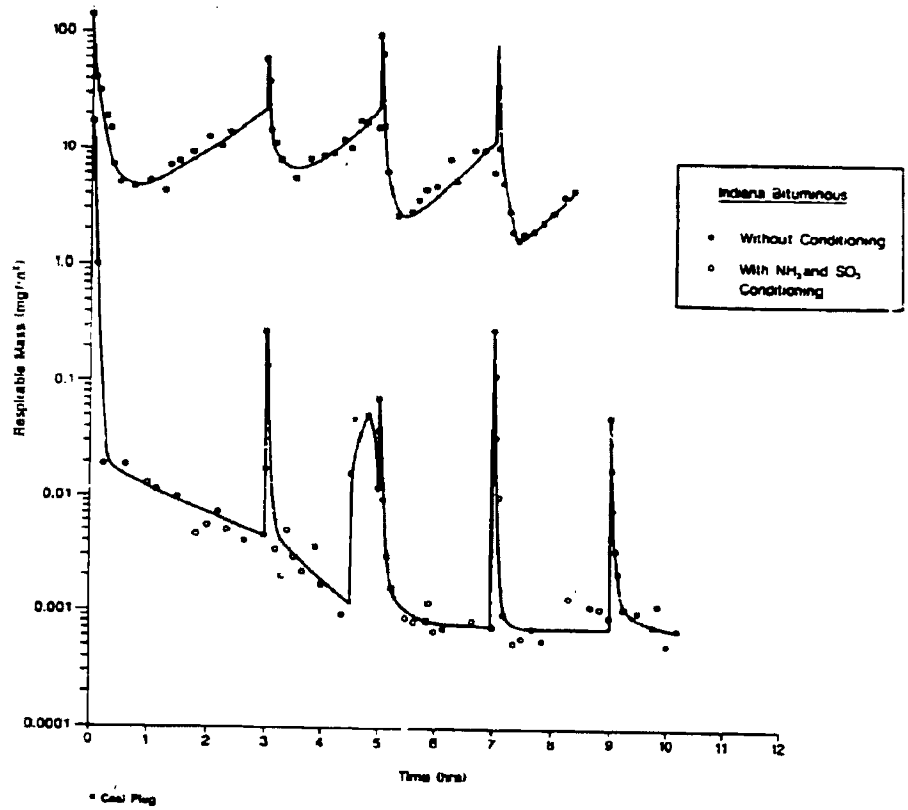

Figure 5. Respirable mass emissions (2 hour shaking interval) for an Indiana bituminous coal.

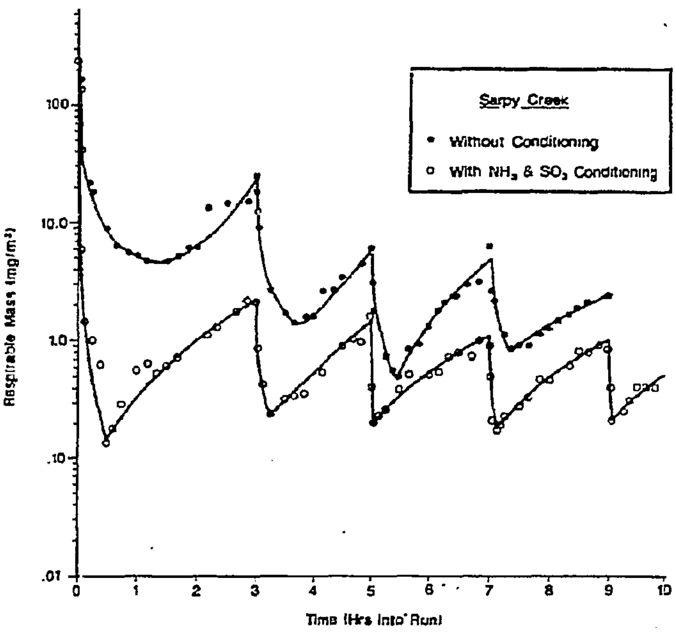

Figure 6. Respirable mass emissions (2 hour shaking interval) with Sarpy Creek subbituminous coal.

Pressure Drop

An important consideration in fabric filtration is the operational pressure drop. The effects of flue gas conditioning on pressure drop must be determined before the potential of this technology for full-scale applications can be evaluated. During the initial dust cake build-up, starting with new fabric, the slope of pressure drop as a function of time was less with flue gas conditioning, indicating that conditioning reduces the dust cake resistance. After initiating the cleaning cycle, the reduction in pressure drop appears to continue. As part of the Phase I tests, to evaluate the longer term effects of conditioning on pressure drop, fifty hour baseline and conditioning runs were completed using South Hallsville lignite. This data is presented in Figure 7. The data points were taken before the cleaning cycle was initiated and immediately before each shake. The pressure drop for the test with SO_3 and ammonia conditioning was lower than for the test performed without conditioning and remained lower throughout the test as is illustrated in Figure 7. This was also true for the short term tests using Big Brown and Falkirk lignites.

For the coals tested during Phase II, a reduction in pressure drop was also seen, as is shown in Figures 8 to 10. It was especially significant for the Indiana bituminous coal. The pressure drop reached 14 inches W.C. at the end of the two hour cleaning cycle without conditioning but only 2.5 inches W.C. with conditioning. Although the other two coals did not show as great an

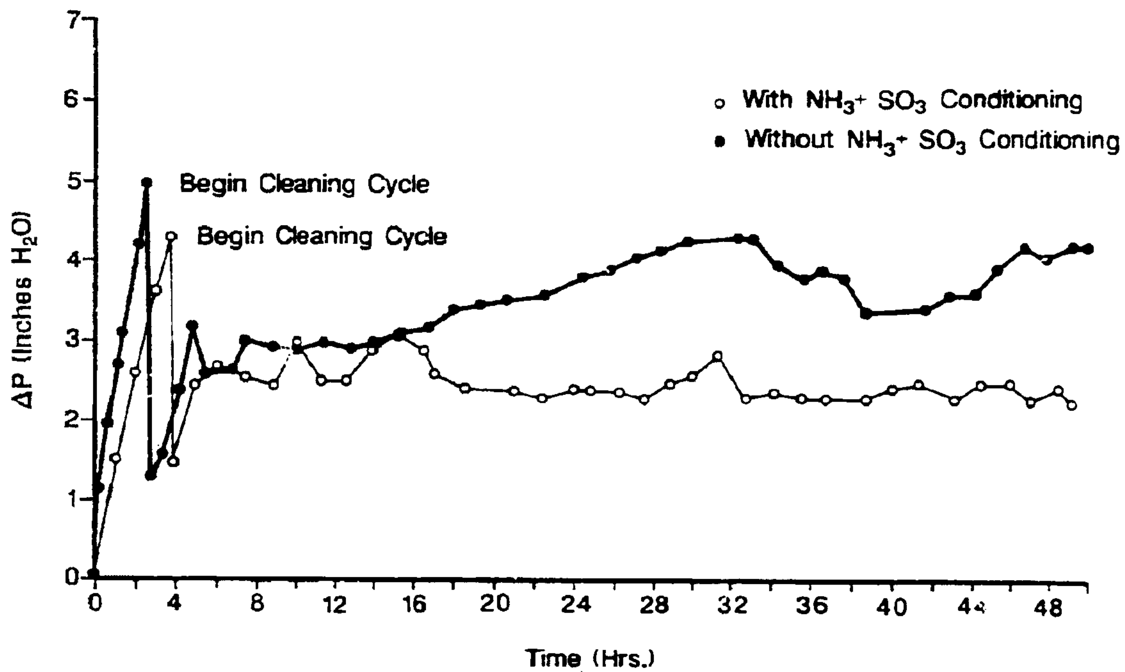

Figure 7. Pressure drop as a function of time for South Hallsville, Texas lignite (50 hour tests).

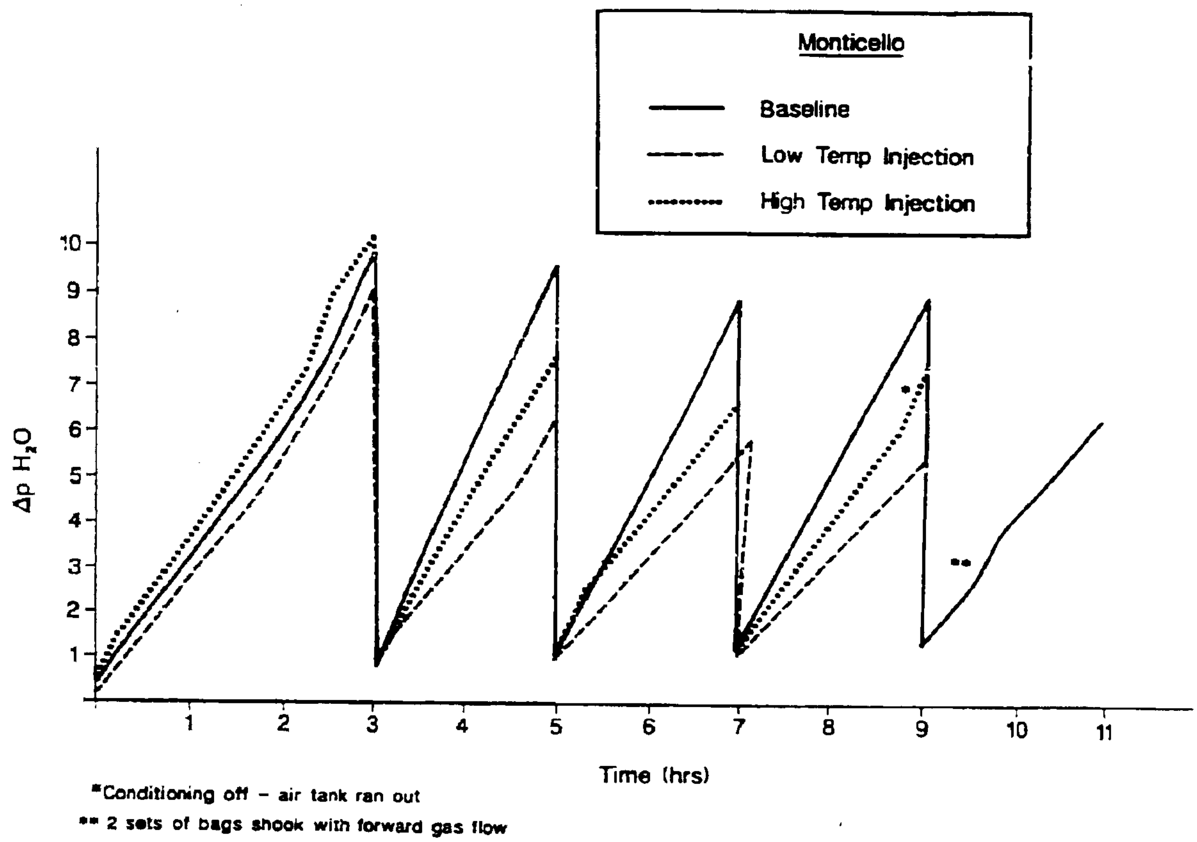

Figure 8. Pressure drop as a function of time for Monticello, Texas lignite.

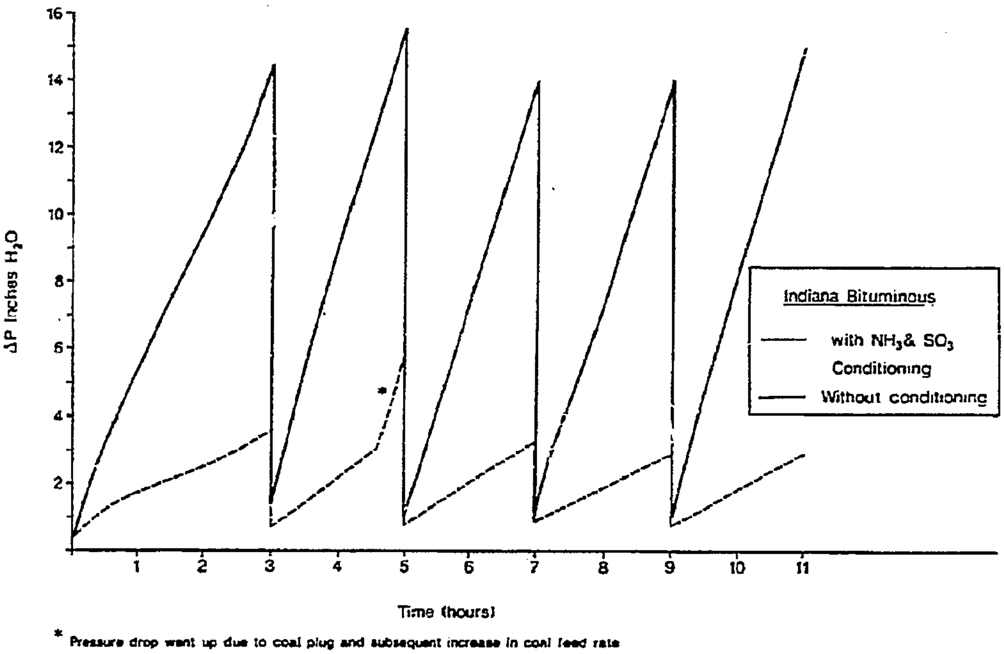

Figure 9. Pressure Drop as a function of time for an Indiana bituminous coal.

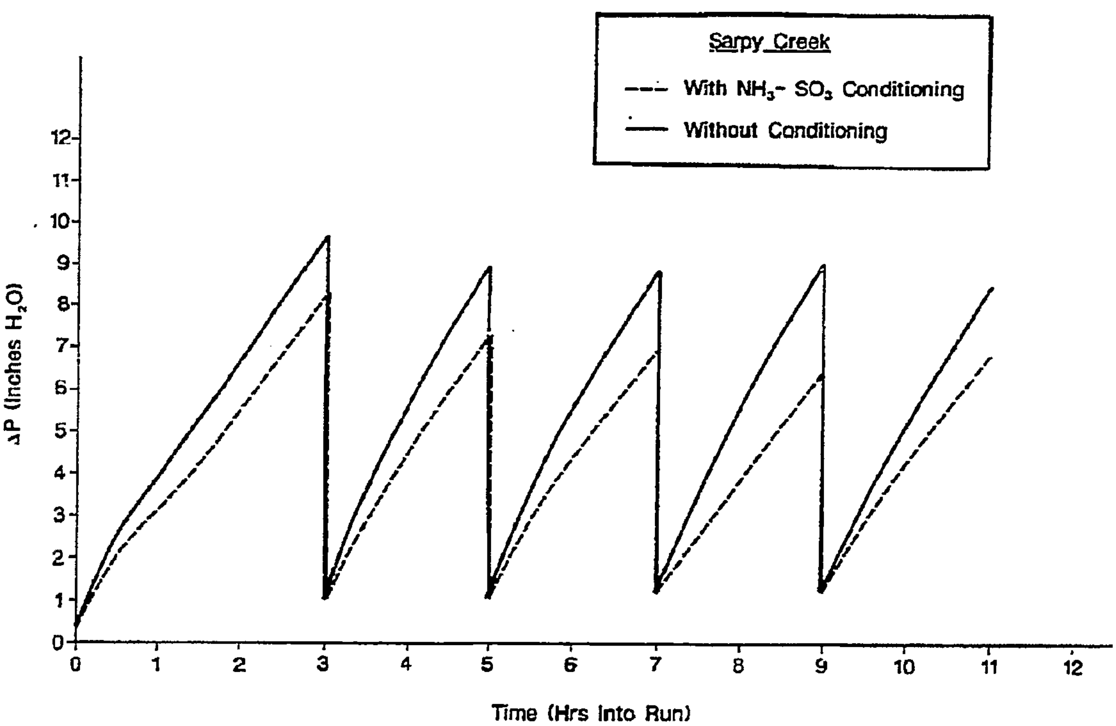

Figure 10. Pressure drop as a function of time for Sarpy Creek subbituminous coal.

improvement, the baghouse pressure drop was still reduced with conditioning. The available data appear to be consistent for all the coals tested, indicating that with SO_3 and ammonia conditioning a more porous dust cake is formed, resulting in a lower pressure drop.

Physical and Chemical Characterization

There must be some physical and chemical changes in the fly ash and/or dust cake to account for the observed enhanced particulate removal and lower pressure drop. Ultimate and proximate analyses for all the coals used in the conditioning tests are given in Table 2. The composition of the baghouse hopper ashes, determined by x-ray fluorescence (XRF) analysis, is given for both the conditioned and nonconditioned tests in Table 3. The important ash component to note from this table is the sodium concentration of each ash. All the coals used in Phase I testing were very low in sodium and all tests without SO_3 and ammonia conditioning resulted in high particulate emissions. In Phase II testing, the sodium concentrations were very low for the Monticello and the Indiana bituminous coals and these coals responded very well to conditioning. The Sarpy Creek subbituminous coal, on the other hand, was quite high in sodium, about 5% sodium as Na_2O. The ash from this coal, as predicted from an ash composition model developed previously (2,3), would collect very well in a baghouse and this was found to be the case. The baghouse efficiency without conditioning was about 99.8% and increased to 99.9% with conditioning. This was not as substantial an improvement as was observed for the other coals. The high sodium concentration may have interfered with the ammonia and SO_3 reactions. For example, the SO_3 may have reacted with the available sodium in the fly ash prior to reacting with ammonia. Leaching the fly ash and analyzing the leachate for ammonia showed that not as much of the ammonia was present on the fly ash as for the other coals. As a result, the ammonia concentration in the flue gas downstream of the baghouse was also higher for the Sarpy Creek test.

The exact ammonia-SO_3 species that is deposited on the fly ash surface is not clear. ESCA (Electron Spectroscopy for Chemical Analysis) has revealed that all the surface sulfur is present as a sulfate rather than a sulfite, which is consistent with results reported by others (7). Being a strong Lewis acid SO_3 will react with the ammonia that is present. At the Phase I injection temperature of 700°F, the likely compound to be formed will be ammonium hydrogen sulfate. Other possible products such as ammonium sulfate, which would be favored at lower temperatures would dissociate at this temperature. Ammonium hydrogen sulfate has a dissociation temperature of 770°F, but dissociation is not complete until 842°F is reached (8). At 700°F ammonium hydrogen sulfate is above its melting point and therefore it must be in a liquid state and in equilibrium with its dissociation products (9). At the Phase II injection temperature of 350°F which is below the ammonium sulfate dissociation temperature, either ammonium sulfate or ammonium hydrogen sulfate could form depending on the concentration of ammonia, SO_3, and other flue gas constituents. The exact chemical mechanisms occurring are complex and more laboratory study of the reactions of ammonia, SO_3, and fly ash at different temperatures is needed.

TABLE 2

ULTIMATE AND PROXIMATE COAL ANALYSES

Coal Analysis (moisture free)	Falkirk	Big Brown	South Hallsville	Monticello	Indiana Bit.	Sarpy Creek
Proximate						
Ash	11.50	18.81	13.71	21.98	11.70	11.31
Volatile Matter	44.81	42.79	39.60	41.24	37.28	38.41
Fixed Carbon	45.89	38.80	46.72	37.78	51.02	50.28
Coal Moisture, % (as received)	36.7	25.3	7.9[a]	34.1	8.7	18.9
Ultimate						
Carbon	61.57	55.60	60.83	56.36	69.47	65.63
Hydrogen	4.26	5.22	3.87	4.51	4.96	4.43
Nitrogen	1.15	0.78	1.32	1.01	1.44	0.83
Sulfur	1.05	1.14	1.28	0.75	2.46	0.73
Oxygen (by diff.)	20.47	18.47	24.52	16.38	9.95	17.05
Heating Value						
Btu/lb	10,515	9,957	10,673	9,753	12,322	11,093

[a] Coal was dried.

Fly ash cohesiveness appears to be an important physical characteristic affecting baghouse performance. Cohesiveness was measured by forming a disk shaped ash pellet (30 mm diameter by 5 mm thickness) in a high pressure press and then applying a force to the center of the pellet while supporting it on its edge. The magnitude of the force required to break the pellet is a measure of the ash cohesiveness. A change in fly ash cohesiveness due to conditioning was observed for all the test coals except for Sarpy Creek, as is shown in Table 4. This is consistent with the baghouse emission results where Sarpy Creek showed the least improvement with conditioning. To help verify this data, samples of the nonconditioned and conditioned Big Brown fly ashes were sent to Southern Research Institute (SoRI) to determine cohesiveness using a shear cell method. This method also indicated a substantial increase in cohesiveness due to conditioning. Other nonconditioned ash samples sent to SoRI showed the same relative cohesiveness as that obtained at UNDERC, confirming the validity of the UNDERC test method.

TABLE 3

XRF ASH ANALYSIS OF BAGHOUSE HOPPER ASH
(Percent Concentration as Oxides)

	SiO_3	Al_2O_3	Fe_2O_3	TiO_2	P_2O_5	CaO	MgO	Na_2O	K_2O	SO_3
Falkirk										
Nonconditioned	42.2	12.1	12.6	0.9	<0.1	20.8	7.0	0.6	1.5	1.8
Conditioned	34.7	12.0	13.2	0.9	<0.1	24.5	8.2	0.9	1.3	4.3
Big Brown										
Nonconditioned	51.5	19.3	8.3	1.7	<0.1	14.4	2.9	<0.5	1.0	1.2
Conditioned	51.9	19.8	7.6	1.7	<0.1	13.7	3.0	<0.5	0.8	1.3
South Hallsville										
Nonconditioned	39.6	12.5	24.4	1.7	<0.1	14.5	2.9	<0.5	0.4	3.2
Conditioned	35.8	12.7	27.5	1.7	<0.1	14.6	3.5	<0.5	0.4	3.5
Monticello (700°F Inject.)										
Nonconditioned	56.1	17.2	4.1	1.9	<0.1	16.6	2.5	<0.5	0.5	1.2
Conditioned	55.2	17.4	4.1	2.0	<0.1	16.8	2.5	<0.5	0.5	1.2
Monticello (350°F Inject.)										
Nonconditioned	56.1	17.2	4.1	1.9	<0.1	16.8	2.5	<0.5	0.5	1.2
Conditioned	55.6	17.1	4.1	2.0	<0.1	17.0	2.5	<0.5	0.5	1.1
Indiana Bituminous										
Nonconditioned	50.5	24.2	13.2	1.4	0.1	3.5	1.7	<0.5	3.6	1.8
Conditioned	50.1	24.4	13.1	1.3	0.1	3.0	1.8	<0.5	3.7	2.5
Sarpy Creek										
Nonconditioned	35.0	20.6	6.4	1.2	0.3	25.5	2.4	5.3	0.7	2.8
Conditioned	34.3	20.9	6.4	1.2	0.3	25.0	2.5	5.0	0.6	3.2

Ash cohesiveness would be expected to affect the residual dust cake weight on the bags. If the ash is too "sticky" it may not be easily removed. On the other hand, if the cohesiveness of the ash is increased while the adhesion to the fabric remains unchanged, bag cleaning may be easier since the dust would be removed in large agglomerates. At the end of the 50-hr tests with South Hallsville and after Phase II tests, bag weights were measured after shaking. The residual dust cake weights for the cleaned bags were reduced up to 50% for the conditioned compared to the baseline tests with, again, Sarpy Creek being the exception. The lower residual dust cake weights seem to indicate that the conditioned ash is more easily removed from the bags, which is consistent with the lower observed pressure drop.

TABLE 4

COHESIVE STRENGTH OF FLY ASHES AS MEASURED BY UNDERC'S COHESIVE TEST METER

Coal Fly Ash	Breaking Force ($lbs \times 10^{-2}$)	
	Noncondltioned	Conditioned
Beulah	52	--
Falkirk	4.4	23
Big Brown	5.5	21
South Hallsville	12	18
Monticello (700°F injection)	10	14
Monticello (350°F injection)	10	18
Indiana bituminous	11	14
Sarpy Creek	15	13

Mechanisms

Pressure drop across the baghouse during initial dust cake deposition can be compared to that in granular bed filter and is directly proportional gas viscosity, superficial gas velocity, and bed depth. The proportionality constant, defined by Darcy's law, is called permeability. An increase in permeability of the dust cake results in a decrease in pressure drop. The gas velocity and viscosity as well as the bed depth would be relatively constant when comparing the conditioned and nonconditioned results. Therefore, only a change in permeability could account for the reduced pressure drop observed with flue gas conditioning.

Kozeny and Carman attempted to relate permeability to other cake properties such as particle size, porosity, and bed structure (10). The particle size distribution and bed structure may be changed somewhat with conditioning but not enough to account for the large change in pressure drop observed. Permeability, on the other hand, is quite sensitive to porosity (void volume fraction); for example, a 20% increase in porosity can cause the permeability to increase by a factor of 2 or more. A probable explanation for the observed reduction in pressure drop is that the conditioned ash forms a more porous dust cake which increases the overall dust cake permeability. A more porous dust cake could result from an increase in ash cohesiveness since it would be less susceptible to packing and pore collapse.

There are two mechanisms, pinhole penetration and seepage, which could account for high particulate emission levels for difficult-to-collect ashes. In pinhole penetration, large pore openings in the weave are not bridged over, resulting in high local gas velocities in comparison to the superficial face velocity (11). Particles which penetrate the fabric through pinholes are never collected. Penetration by seepage refers to particles that were initially collected but eventually seep through the fabric (12). Some seepage is expected during cleaning, if pinhole plugs break loose, or if ash is reentrained. An increase in cohesiveness due to flue gas conditioning would facilitate bridging large pores and thus reduce pinhole penetration. In addition, a more cohesive fly ash would minimize penetration by seepage since particles collected would be held more firmly in place.

ECONOMICS

The technical merits of flue gas conditioning are very strong; however, it must also be economically viable to be considered as an enhancement method for particulate control. A preliminary economic evaluation was completed to determine the cost of flue gas conditioning compared to conventional fabric filtration. The design criteria (13) used to determine the size and overall cost of a baghouse for this economic study include the following:

1. The plant is 500 MW with a pc-fired boiler.
2. Air-to-cloth ratio is 2.0 ft/min.
3. The baghouse has 16 compartments with 744 bags per compartment.
4. Bags are woven fiberglass with a Teflon B finish and have a 4 year bag life with a 25% annual replacement rate.
5. The baghouse is operated at a nominal temperature of 300 °F and is located on the negative side of the ID fan.
6. Desired maximum operating tube sheet pressure drop is 4.5 inches W.C..
7. The cleaning mechanism is reverse air with a 1 hour cleaning cycle.
8. Inlet dust loading is 3 to 4 grains/scf.
9. Desired collection efficiency is 99.9%.

The total capital investment (TCI) for the fabric filtration system using these design criteria is $23,364,000 or $44.73/KW. The annual levelized cost is $11,207,800/yr or 3.01 mills/KWh. These costs were obtained from an Electric Power Research Institute report and up-dated to 2nd quarter 1986 dollars (13). These were the base costs to which flue gas conditioning costs were compared.

The economics of conditioning flue gas with ammonia and SO_3 will depend on many factors. Some of these, such as quantity of agents needed, the effects of coal type, and the effects of temperature, still need to be determined experimentally. For this economic evaluation an injection rate of 25 ppm ammonia and 12 ppm SO_3 was assumed. In addition, it was assumed that conditioning will not adversely affect ash handling or bag life. Capital and operating costs are determined based on air-to-cloth ratios of 2 and 2.7 ft/min. The cost of the conditioning equipment, as well as operating costs, were obtained from Wahlco Inc.. Capital costs include the cost of equipment, installation, shipping, and supervision, as well as a contingency fund to provide for unforeseen events. Operating costs for conditioning include the costs of sulfur and ammonia, additional energy needed (electrical and steam), maintenance, and labor. Costs are summarized in Tables 5 and 6. For the economic evaluation, the gas flow and the number of compartments were assumed to remain the same but with a reduction in the total number of bags. If the injection of ammonia and SO_3 can reduce pressure drop, such that the baghouse can operate at an air-to-cloth ratio of 2.7 ft/min, then the TCI will be reduced by 10.8%. The annual levelized cost savings at an air-to-cloth ratio of 2.7 to 1 would be 9.3%. The annual levelized cost break even point, comparing fabric filters with and without conditioning, would occur for a fabric filter with conditioning operating at an air-to-cloth ratio of 2.32 ft/min.

TABLE 5

COMPARISON OF CAPITAL COSTS FOR FABRIC FILTRATION SYSTEMS WITH AND WITHOUT FLUE GAS CONDITIONING FOR A 500 MW PLANT UNITS $1000 (1986)

Cost Items	Conventional[a] A/C = 2:1	Conditioned A/C = 2:1	Conditioned A/C = 2.7:1
Turnkey Cost Fabric Filter	22,364	22,364	17,444
SO_3 Generator plus Supports		1,300	1,300
Ammonia Injection Equipment plus Supports		260	260
Freight		8	8
Installation		515	515
Engineering		208	208
Contingency		208	208
Total Capital Investment	22,364	24,863	19,943

[a] Reference 13.

Table 6

COMPARISON OF OPERATING AND MAINTENANCE COSTS FOR A FABRIC FILTRATION SYSTEM WITH AND WITHOUT CONDITIONING
UNITS $1000 (1986)

Cost Items	Conventional[a] A/C = 2:1	Conditioned A/C = 2:1	Conditioned A/C = 2.7:1
Fabric filter maintenance and bag replacement	2,187.7	2,187.7	1,895.0
Ash Removal	450.7	450.7	390.4
Conditioning			
Sulfur @ $175/tn del.		57.4	57.4
Ammonia @ $165/tn del.		60.7	60.7
Miscellaneous		5.0	5.0
Labor	709.6	867.3	690.6
Energy Cost	2,492.4	2,661.7	2,256.9
Amortization	5,367.4	5,967.1	4,786.3
Levelized Annual Operating Cost	11,207.8	12,257.6	10,142.3
Mills/KWh	3.01	3.27	2.73

[a] Reference 13.

The future intentions of government regulatory agencies will have an impact on the potential of flue gas conditioning technology. If particulate control regulations become more stringent and if fine particulate standards are enacted, then flue gas conditioning may well become the economic and technological choice.

CONCLUSIONS

Although this study was exploratory in nature and presents relatively new data, several conclusions can be made:

- Results clearly show enhanced particulate removal with SO_3 and ammonia conditioning for coals that produce difficult-to-collect ash, for a given fabric and cleaning cycle.

- For the given coals and test conditions, conditioning reduces pressure drop both during initial dust cake deposition and after initiation of the cleaning cycle.

- Although the physical and chemical mechanisms are complex, surface deposition of an ammonia-SO_3 product and a resulting increase in fly ash cohesiveness partially explain the higher collection efficiency and lower pressure drop.

- The conditioning effect does not appear to be dependent on the injection temperature and therefore the injection point may be after the air preheater.

- With all else being equal, the economics indicate that to break even with conventional baghouses, the air-to-cloth ratio would need to be 2.32 to 1 using conditioning compared to 2.0 to 1 without conditioning.

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CONVERSION OF SAN ANTONIO CITY PUBLIC SERVICE J.T. DEELY UNIT NO. 2 ELECTROSTATIC PRECIPITATORS FROM HOT SIDE TO COLD SIDE OPERATION

by J.E. Pruske and J.R. Schwegmann
City Public Service, San Antonio, Texas

The J.T. Deely Power Plant of San Antonio City Public Service consists of two (2) 418-MW western coal-fired units. J.T. Deely Unit #1 went into commercial operation on August 8, 1977 and Unit #2 on August 1, 1978.

Each unit consists of a Combustion Engineering, Inc. balanced draft, drum type steam generator, a single trisector Ljungstrom air preheater, and two Buell hot side, weighted wire precipitators. The decision to use hot side weighted wire precipitators for flue gas particulate removal was based on available technology and utility/consultant thinking during the early to middle 1970's.

Soon after the units went into operation in 1977 and 1978, CPS began to experience the well publicized problems associated with hot side precipitators. In general, the problem can be characterized by a gradual deterioration of performance. This deterioration was caused by the combined effects of ash buildup on the discharge electrodes, development of a sodium depleted ash layer on the collecting plates, poor plate to wire clearances due to thermal expansions and contractions, and inadequate wire cleaning. After about 10 to 14 weeks of high load operation, the emissions from the precipitators would exceed opacity limitations. This forced CPS to shutdown the coal unit and thoroughly wash each precipitator. The J.T. Deely coal units are the most economical generating units on the City Public Service system.

As the years went by, the frequency of washing the precipitators increased. These frequent shut downs for a precipitator wash not only affected the integrity of the precipitators but also the performance of the entire coal unit. The inability of the J.T. Deely hot side precipitator to maintain satisfactory performance became unacceptable to CPS. This paper summarizes the studies and work performed to convert the precipitators from hot side to cold side operation.

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INTRODUCTION

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INITIAL STUDIES

The concept of converting the precipitators from hot side to cold side operation was first presented to City Public Service in 1984 by Wahlco, Inc. a California based company. Wahlco studied the hot to cold conversion concept for several months and in May, 1985 presented City Public Service with their findings on the precipitator conversion concept. Wahlco concluded that by redirecting the flue gas flow to the air preheater before entering the precipitators and conditioning the flue gases with sulfur trioxide, precipitator performance would be improved and the units would remain in compliance with the 20% opacity requirement. Wahlco also concluded that boiler efficiency would be improved primarily as a result of the reduction in air in-leakage upstream of the air preheater. The estimated cost for the conversion of both J.T. Deely units precipitators to cold side operation was $21,000,000.

Before Wahlco's conversion idea came along, CPS had been considering the replacement of the hot side precipitators with new cold side precipitators or baghouses. However, the $100 million cost for this plan was considered extremely expensive by CPS. Therefore, CPS began to pursue the precipitator conversion concept in more detail. Specific areas of concern to CPS engineering and operations personnel were:

1. The structural integrity of the existing hot side precipitators.
2. The adaptability of the existing trisector air preheater to handle the ash laden flue gases leaving the economizer.
3. The adaptability of the existing precipitators to cold side operation.
4. The effect of sulfur trioxide conditioning on flue gas emissions from the plant.

To answer these questions, CPS began our own investigation into the precipitator conversion concept. These investigations are summarized below:

Precipitator Structural Integrity

CPS retained Ebasco Services, Inc. to conduct an independent structural investigation of the hot side precipitators. Ebasco's investigation involved a review of the Buell precipitator drawings and Black & Veatch ductwork drawings, a field inspection of the J.T. Deely Unit #1 precipitators, and a discussion with CPS personnel to obtain an operational history of the precipitators. Ebasco concluded that the structural design and the existing steel components of the precipitators are adequate for the existing hot side service and even more so for the lower operating temperatures expected after conversion to cold side. Ebasco made several recommendations for improvements/repairs to the existing precipitator systems.

Feasibility Study of Precipitator Conversion

CPS retained Environmental Management Associates (EMA) to perform a technical feasibility study for the conversion of the precipitators from hot side to cold side operation and recommend to CPS whether the proposed conversion is viable. EMA performed a very detailed and timely analysis and concluded that the conversion of the hot side precipitators to cold side operation is technically feasible and should result in high reliability operation and in continuous long term compliances with air pollution regulations. EMA recommended several design changes/modifications necessary to achieve this desired result. These recommendations included:

- Straighten all bowed and warped plates
- Repair, replace or redesign (if necessary) the plate to plate clips and mid-point guide rakes to maintain plate alignment
- Redesign the lower guide system to insure free downward expansion of the plates while still retaining proper alignment
- Substitute compression insulators for the tension insulators
- Install three vibrators/rappers per electrical bus section in lieu of one
- Level and realign the discharge electrode system
- Repair and/or replace the precipitator roof insulation
- Perform model study and incorporate recommended gas distribution devices
- Replace access door seals to reduce air in-leakage
- Change separate insulator compartments to insulator houses
- Install air purge and heater system for insulator houses
- Install new hopper heating system over bottom half of hoppers and throat
- Replace the hopper level alarms with a nuclear type level alarm
- Consider modifying the air preheater to reduce the temperature spread and average flue gas temperature leaving air preheater

Air Preheater Investigations

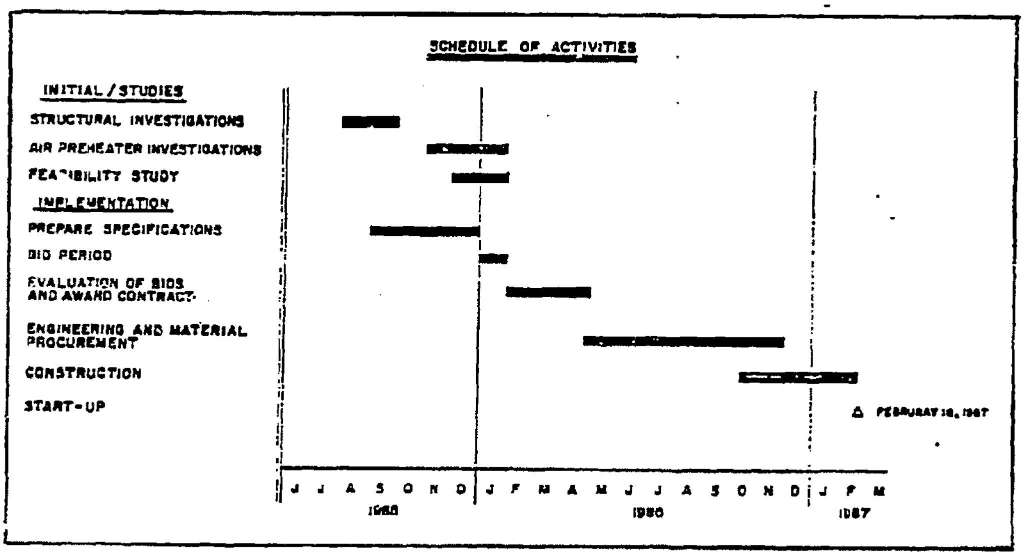
CPS met with Combustion Engineering Air Preheater on several occasions to discuss the adaptability of the existing trisector Ljungstrom air preheater to cold side precipitator operation. Combustion Engineering Air Preheater recommended that the existing air preheater baskets be replaced with a loosely packed basketed element which would improve the cleanability of the baskets by soot blowing. Also, Combustion Engineering Air Preheater determined that an additional six inches of basket depth could be added to further reduce the temperature of the flue gas leaving the air preheater. Combustion Engineering Air Preheater provided preliminary performance and cost information which was not considered a fatal flaw to the precipitator conversion concept.

Sulfur Trioxide (SO_3) Emissions

If flue gas conditioning with SO_3 was required to meet opacity requirements, CPS was concerned about the impact that SO_3 injection would have on SO_3 emissions from the unit. Based on conversations with utilities with flue gas conditioning systems and based on several articles which have been written on SO_3 flue gas conditioning, SO_3 is readily absorbed by the flue ash particles at the lower flue gas temperatures and thus no additional SO_3 emissions should be expected. In fact, the lower temperature of the flue gas following conversion to cold side precipitation should enhance free SO_3 removal.

IMPLEMENTATION

Concurrent with the above studies and investigations, CPS began to prepare the necessary specifications and documents for the precipitator conversion project. Every effort was made to perform the first precipitator conversion (J.T. Deely Unit #2) during the fall and winter of 1986-87 in order to coincide with a planned major rework of the turbine's low pressure spindle. A schedule of the activities for the J.T. Deely Unit #2 precipitator conversion is shown below:



CPS wrote the specifications and documents in such a manner so as to place full responsibility for precipitator performance upon a single source. CPS required each prospective bidder to be the lead company in a partnership which must include a general contractor, a architect-engineer, a precipitator design company, a SO_3 flue gas conditioning system company, and a flow modeling company.

The successful bidder, or contractor was required to guarantee that the converted precipitators would:

1. Maintain opacity at or below 20%, and
2. Maintain mass emissions at or below 0.1 lb/MMBtu, and
3. Maintain the precipitator collection efficiency at or above 99.4%.

Compliance with this guarantee is to be demonstrated through everyday readings from the opacity monitors and through actual field testing. The guarantee period is for one (1) year following initial operation.

Also, the successful bidder, or contractor was required to perform the actual conversion work on J.T. Deely Unit #2 during a ten (10) week overhaul outage scheduled for December 6, 1986 through February 16, 1987. Only six (6) weeks was allowed for the J.T. Deely Unit #1 conversion.

The contractors scope of work included:

- Perform model study to determine what modifications are required to the flow distribution devices and ductwork so as to minimize pressure losses and achieve good flue gas flow distribution in the converted precipitators.
- Perform draft study, air preheater study and I.D. Fan study to define the new operating conditions for the J.T. Deely units, to determine the adaptability of the existing equipment to cold side precipitator operation, and to recommend changes/modifications as required.
- Remove existing ductwork and design and install new ductwork as required for cold side precipitator operation.
- Modify existing structural steel so as to support the new ductwork.
- Provide and install an SO_3 flue gas conditioning system.
- Modify existing precipitators for cold side precipitator operation as discussed later in this report.

CPS retained the right to review and approve all drawings and equipment and inspect the work.

Nine companies were requested to bid. Four companies submitted bids. CPS evaluated the bids and issued a contract to H.B. Zachry Company on April 17, 1986 for the J.T. Deely Unit #2 precipitator conversion.

SCOPE OF WORK

Below is a detailed discussion of the work performed to convert the J.T. Deely Unit #2 precipitators from hot side to cold side operation with flue gas conditioning.

Performance Study

One of the first tasks of the Contractor was to define the new operating conditions for the J.T. Deely units with a cold side precipitator conversion. Of specific concern was the adequacy of the existing air preheater and induced draft fans to perform continuously at full load conditions with a cold side precipitator conversion. Combustion Engineering performed the studies and determined that the precipitator conversion will have the following effects on unit performance:

1. The relocation of the precipitators downstream of the air preheater will eliminate the precipitator heat loss and air in-leakage effects on air preheater and boiler performance. With the flue gases going directly from the economizer to the air preheater more heat will be returned to the boiler through the air preheater thus, increasing boiler efficiency.

2. With the existing hot side precipitator arrangement, sodium was added to the coal to improve precipitator performance. Also, economizer soot blowing was limited in order to maintain a high flue gas temperature entering the hot side precipitators so as to maintain precipitator performance. Both the sodium and limited soot blowing, compounded by frequent shutdowns to wash the precipitators, resulted in severe economizer plugging. This pluggage resulted in a high draft loss across the economizer which limited I.D. fan and boiler capability. The cold side precipitator conversion will eliminate the need for sodium, will allow for frequent economizer soot blowing and thus, reduce the draft loss across the economizer and increase the capability of the boiler.

Table 1 shows the boilers performance with the existing hot side precipitators and the predicted boiler performance with a cold side precipitator.

PREDICTED BOILER PERFORMANCE

	HOT SIDE ESP EXISTING-80" TIGHT-PACKED ELEMENTS	COLD SIDE ESP 80" LOOSELY PACKED APH ELEMENTS(1)	COLD SIDE ESP 86" LOOSELY PACKED APH ELEMENTS(1)
MAIN STEAM - LBS/HR	3,061,000	3,061,000	3,061,000
HEAT OUTPUT - 10^6 BTU/HR	3,579	3,565	3,565
EXCESS OXYGEN - %(2)	4.65	4.0	4.0
BOILER EFFICIENCY - %(3)	83.15	83.86	84.10
HEAT FIRED - 10^6 BTU/HR	4,304	4,251	4,239
FUEL HHV - BTU/LB	8,156	8,156	8,156
FUEL FIRED - LBS/HR	527,700	521,200	519,700
APH FLOWS - LBS/HR			
GAS ENTERING	4,644,000	4,434,000	4,421,000
PRI. AIR BYPASS (4)	145,000	133,000	133,000
PRI. AIR LEAVING (4)	760,000	760,000	760,000
SEC. AIR LEAVING	2,967,000	2,962,000	2,951,000
GAS LEAVING (CORR'D FOR LEAKAGE)	5,014,000	4,781,000	4,770,000
APH TEMPS. - °F			
PRI. AIR ENTERING	90	90	90
SEC. AIR ENTERING	85	85	85
GAS ENTERING	820	810	810
PRI. AIR LEAVING	761	746	752
SEC. AIR LEAVING	738	721	731
GAS LEAVING (UNCORR'D)	340	320	314
GAS LEAVING (CORR'D FOR LEAKAGE)	323	305	298

NOTES:
1. AIR PREHEATER IN AS-NEW CONDITION.
2. AT AIR PREHEATER INLET.
3. BASED ON EXCESS AIR AT AIR PREHEATER EXIT GAS TEMPERATURE (INCLUDES 1.5% UNACCOUNTABLE LOSS).
4. ESTIMATED VALUES.

TABLE 1

Because of the lignitic ash characteristics of the fuel being fired and the anticipated increased ash loading to the air preheaters with the precipitator conversion, Combustion Engineering recommended that the existing hot end and intermediate layers of element be replaced with a loosely packed basketed element. This loose pack element will allow somewhat larger ash particles to pass through the air preheater and improve the cleanability of the basketed elements by soot blowing. In addition, Combustion Engineering recommended that an ash deflector be installed above the economizer gas outlet duct in order to reduce the carry over of any large ash particles to the air preheater.

Because cold side precipitator performance is affected by the flue gas temperature and temperature swing leaving the air preheater, Combustion Engineering was asked to identify what modification could be made to reduce the flue gas temperature and temperature swing leaving the air preheater. Combustion Engineering determined that the flue gas temperature could be reduced below 300° F by increasing the element depth from 80 to 86 inches. This could be accomplished in a three-layer configuration without casing modifications or structural modifications to the air preheater rotor. In order to reduce the temperature swing leaving the air preheater from about 60° F to 31° F Combustion Engineering recommended that the speed of the rotor be increased by approximately 20%. This modification required a lower ratio gear reducer, a new electric drive motor and an extra auxiliary air motor.

__Model Study__

Another of the early tasks of the Contractor was the construction of a 1/12 scale model in order to establish the gas flow performance of the precipitator and ductwork from the economizer outlet to the I.D. fan inlets to evaluate draft losses, ash fallout, and gas distribution through the precipitator before and after the conversion. Nels Consulting Services constructed the model and performed the model studies.

Because of the lower flue gas temperature, flue gas velocities through the converted cold side precipitator are much lower than for the existing hot side arrangement. Due to these lower flue gas velocities, Nels recommended several modifications to the precipitator inlet nozzle and outlet plenum flow distribution devices in order to improve flue gas flow distribution through the precipitators. In addition, Nels recommended that the air jet system (fluffing air) in the bottom of the precipitator inlet nozzles be put back into service in order to remove expected ash deposits in this location. Nels predicted that the pressure loss for the total system would be 1.6 inches less than the existing system based on the expected gas flow conditions.

The above recommendations were incorporated into the precipitator conversion work by the Contractor without affecting the schedule.

Ductwork/Structural Modification

In order to convert the Precipitators to cold side operation, the Contractor was required to:

1. Remove all existing ductwork from the economizer outlet to the I.D. fan inlet.
2. Design and install new ductwork from the economizer outlet to the air preheater, from the air preheater to the precipitator inlet, and from the precipitator outlet plenum to the I.D. fans.
3. Modify existing structural steel as required to provide space for the new ductwork and to support the new ductwork.
4. Design and install new ash hoppers under the vertical section of ductwork from the air preheater outlet to the precipitator inlet.

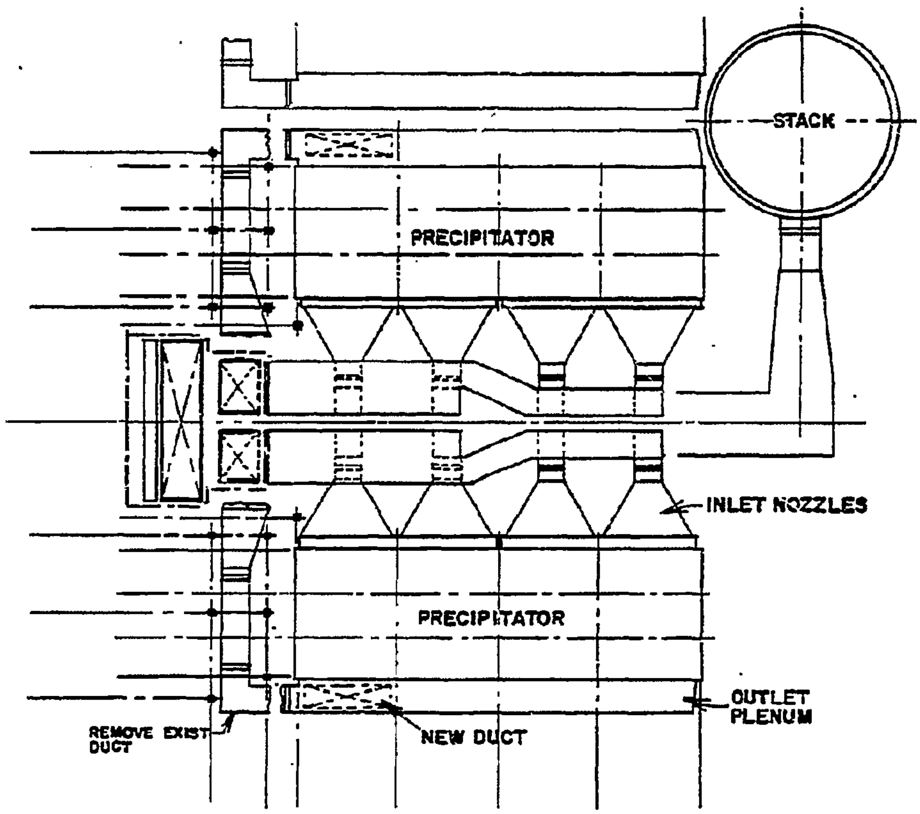
Figure 1 shows the ductwork arrangement for the conversion to cold side precipitator operation. CPS was fortunate in that the existing plant arrangement allowed for this ductwork change without having to relocate any major equipment.

Precipitator Modifications

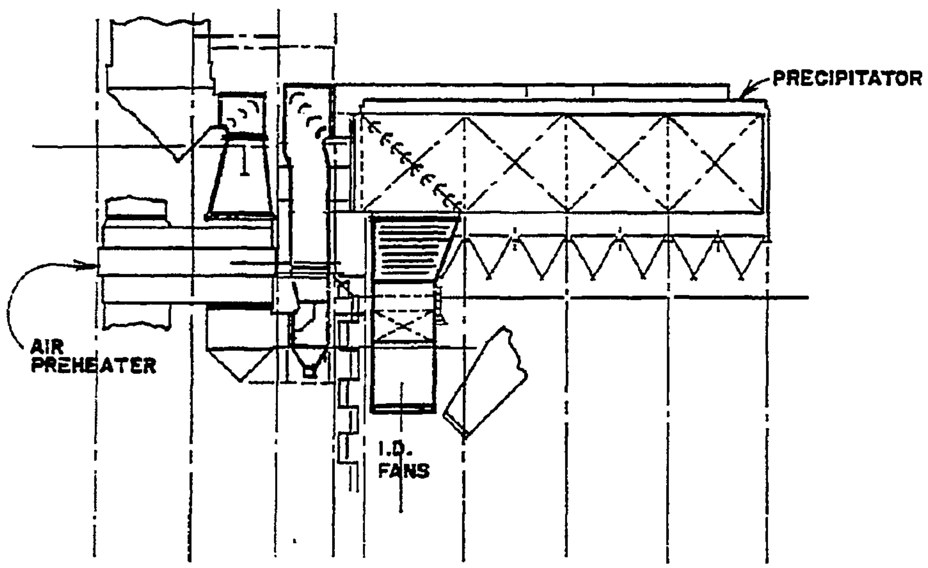
The Contractor was required to make several modifications to the existing precipitators to allow for reliable long term performance. These modifications included:

1. Rework of precipitator internals necessary to obtain the alignments, clearances, and plumbness of emitting wires and plates such that performance guarantees are met. This included the removal of collecting plates which could not be straightened, leveling of rapper/collection plate beams, repairing/replacing plate to plate clips, and redesign of midpoint guide rakes.
2. Replace all tension insulators with compression insulators.
3. Replace the existing discharge electrode vibrator with a rapper and install two additional rappers per discharge electrode emitting box beam.
4. Install a new heater/blower system and ducting to purge insulators.
5. Remove existing precipitator roof insulation and install new insulation and weather tight checkered plate walking surface above the insulation.
6. Replace all precipitator access doors with new leak-tight cast iron doors.

FIGURE 1



PLAN



SIDE ELEVATION

7. Install new "state of the art" automatic voltage control system.
8. Install new blanket type hopper heaters on lower half of all hoppers.
9. Seal weld cracks in precipitator roof and casing to minimize air in-leakage.

Table 2 shows a comparison of the hot side and cold side precipitator design information.

Air Preheater Modification

As discussed in the feasibility study performed by Environmental Management Associates, the temperature of the flue gases entering the precipitator has a significant effect on precipitator performance. Reducing the flue gas temperature entering the precipitators results in better precipitator performance and less need for SO_3 flue gas conditioning.

In order to reduce the average flue gas temperature entering the precipitators to below 300° F, CPS chose to modify the air preheaters as recommended by Combustion Engineering. This modification involved:

1. Replacing all existing air preheater baskets with loose pack basketed elements.
2. Increasing the basketed element depth from 80 inches to 86 inches by changing the air preheater basket arrangement from a four-layer configuration to a three-layer configuration.
3. Rebuilding the existing Falk speed reducer and installing a larger drive motor and auxiliary drive system in order to increase the speed of the air preheater and thus reduce the temperature swing leaving the air preheater to +/- 31° F.

CPS purchased the materials necessary for the above modifications and hired a separate contractor to replace the baskets and change the air preheater drive system.

Flue Gas Conditioning System

As required by the specifications, the Contractor furnished and installed a flue gas conditioning system for injecting up to twenty (20) ppmv SO_3 into the flue gas stream entering the precipitators. The system furnished is a Wahlco sulfur burning system.

The purpose of the SO_3 flue gas conditioning system is to inject trace amounts of SO_3 into the flue gas stream ahead of a cold side precipitator so as to bring the resistivity of the fly ash to a level at which an electrostatic precipitator can function efficiently. Many utilities have installed SO_3 flue gas

PRECIPITATOR COMPARISON

	J. T. DEELY UNIT #2	
	HOT SIDE ESP'S	COLD SIDE ESP'S
Plant Megawatts	418	418
Precipitator Collection Efficiency	<99.4%	99.4%+
Gas Volume (ACFM)	2,715,000	1,600,000 (Max)
Gas Temp (Precipitator Inlet °F)	800	290
No. of Precipitators	2	2
Total Precipitator Collection Area Ft	850,176	850,176
Specific Collection Area (Ft /1000 ACFM)	313	531.4
Drift Velocity (CM/sec)	8.3	4.9 (Min)
No. of Fields in series (1)	2 @ 9 ft 3 @ 6 ft	3 @ 6 ft 2 @ 9 ft
No. of Chambers per Precipitator	4	4
No. of Gas Passages per Chamber	41	41
No. of TR Sets per Precipitator	20	20
Size TR Sets per Precipitator	(16) 1400 MA @ 45KV and (4) 1800 MA @ 45KV	(16) 1400 MA @ 45KV and (4) 1800 MA @ 45KV
Spacing of Gas Passages	9"	9"
Collection Plate Height	36 ft	36 ft
Electrode Rappers	48 Vibrators	144 Rappers
Treatment Time (secs)	7.17	12
Average Gas Velocity ft/sec	5.02	3.01
Opacity (without flue gas conditioning)	20%+	20%
(with flue gas conditioning)		12%
Inlet Grain Loading GR/SCF	1.87-3.82	1.87-3.82
Service	Base	Cycling

TABLE 2

conditioning systems in order to improve precipitator performance.

The process for generating SO_3 for flue gas conditioning consists of a system comprised of a molten sulfur storage tank, a sulfur metering pump, a sulfur burner and SO_2 to SO_3 converter, and an injection system. Liquid sulfur is pumped to the burner/converter where the sulfur is burned to form SO_2 and converted to SO_3 by means of a vanadium pentoxide catalyst. The SO_3 /air mixture from the converter is injected into the flue gas duct through a distribution manifold and injection probes.

CONSTRUCTION/TESTING

Construction work for the J.T. Deely Unit #2 precipitator conversion started on October 6, 1986, nine weeks ahead of the planned start date. This early start was brought about by the additional time needed to send the turbine low pressure spindle to Westinghouse in order to incorporate the necessary modifications recommended by Westinghouse. These first nine weeks were primarily devoted to tearing down ductwork, removing precipitator roof insulation and equipment, and performing repair work. This repair work included:

- Modifications to the economizer outlet duct which included the replacement of cracked and corroded structural members and repairing all duct plate tears to prevent air in-leakage.
- Modifications to the precipitator roof and roof mounted collecting electrode support hanger assemblies which had become deformed due to long term operation at high temperatures.
- Repair of numerous cracks and tears in the precipitator casing, inlet nozzles and outlet plenum in order to reduce air in-leakage.
- Replacement of the lower collecting electrode spacer bar due to poor design.

Modifications to structural steel, installation of new ductwork and rework of the precipitators commenced as materials were received on the jobsite. Material delivery was a problem in performing the precipitator conversion work during the scheduled outage period; however, the Contractor worked ten hours a day, seven days a week plus a second ten hour shift, when required, to meet the scheduled February 15, 1987 start-up.

Throughout the duration of the precipitator conversion work, CPS inspectors inspected the Contractor's work and worked with the Contractor to solve any problems or deficiencies that were identified. Weekly construction meetings were held with the Contractor to discuss the progress of the work and any problems which needed quick response.

As a quality assurance check to ascertain that minimum acceptable clearances had been attained between collection plates and wires in the precipitator, the Contractor was required to perform an Air Load Test and obtain an electrical reading of 42kV on all cabinets with all wires and plates in place. The Contractor was able to obtain the 42kV reading on all cabinets in time to meet the scheduled start-up date.

In addition to the above, CPS pressurized the ductwork and precipitators with air to five inches of water pressure to check for air tightness. Several small leaks were identified and repaired prior to startup of the unit.

OPERATIONAL STATUS/CONCLUSIONS

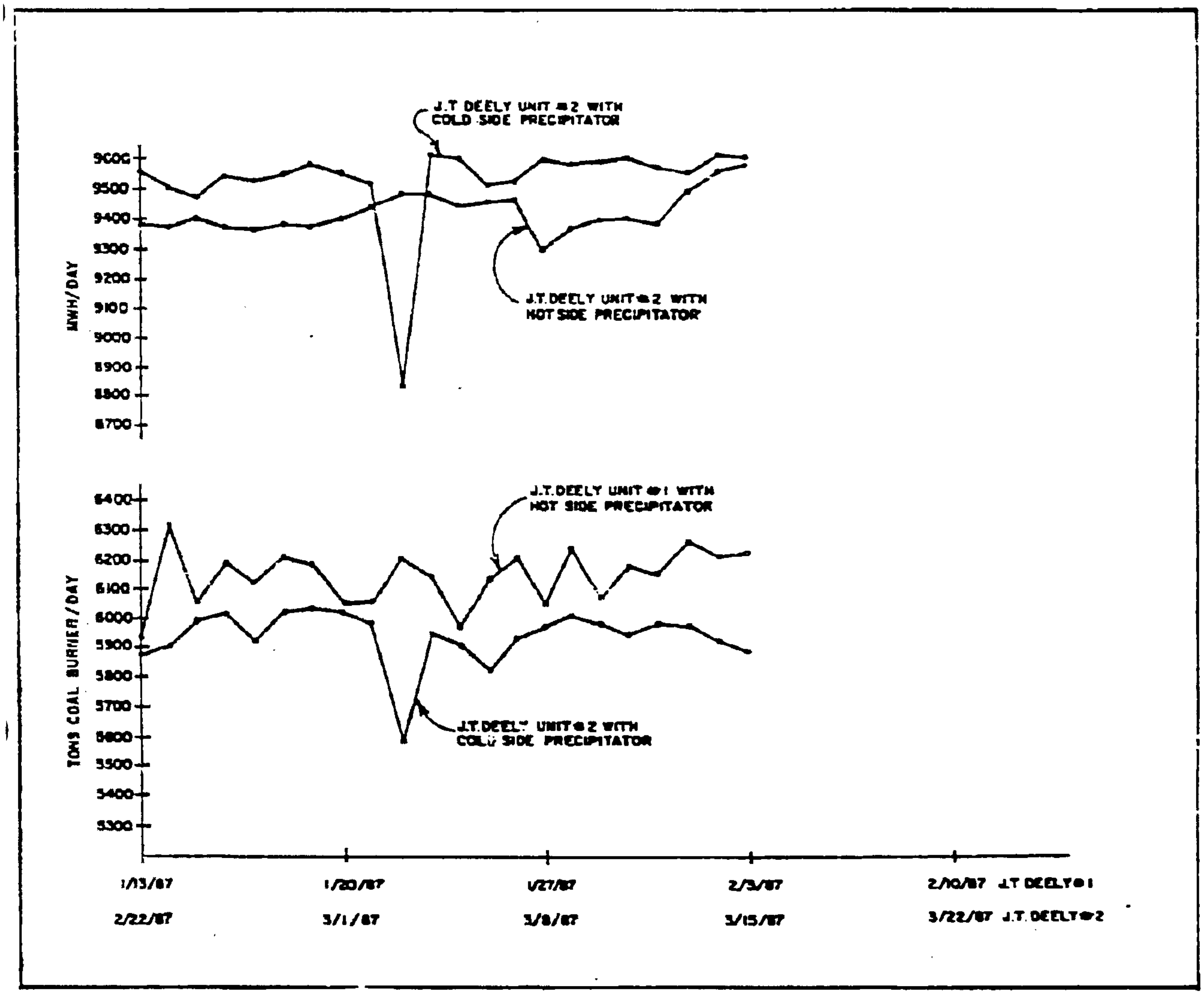
On February 16, 1987, the boiler was fired with fuel oil in order to warm up the boiler and roll and soak the turbine prior to placing the unit in service. On February 17, 1986, the unit was synchronized with the system and began operation on coal. Initial operation has been very good and opacity has remained below 20% for the first full month of operation. Flue gas conditioning has not been required.

Initial operation has shown that A-side precipitator has a much lower opacity than B-side precipitator, primarily due to the 30° F difference in flue gas temperature. The opacity on A-side has ranged between 6-12% while the opacity on B-side has ranged between 12-25%. As discussed above, the temperature difference is due to the temperature swing leaving the air preheater.

Initial operation and testing of the flue gas conditioning system has resulted in a significant improvement in opacity when injecting only a small amount of SO_3 into the flue gas.

One of the major benefits of the precipitator conversion project has been the improvement in boiler efficiency and unit heat rate. As a result of the relocation of the precipitators outside the boiler envelope, the modifications to the air preheater, and repairing of numerous tears and cracks to reduce air in-leakage, more heat is being returned to the boiler thus, improving boiler efficiency. In addition, the draft loss through the system has been reduced to a point where the two-speed I.D. fans can operate at the low speed when carrying full load on the unit. Before the conversion, the I.D. fans were often the limiting factor in carrying full load on the unit.

The chart on the next page shows a comparison of J.T. Deely Unit #1 with a hot side precipitator and J.T. Deely Unit #2 with a cold side precipitator under base load conditions. As can be seen from the chart, J.T. Deely Unit #2 with a cold side precipitator is generating more electricity and burning less coal than J.T. Deely Unit #1. This will result in significant savings to CPS.



The initial operating and performance of the converted precipitators has been very good; however, the true test is how well the unit will perform in the months and years ahead.

FABRIC AND FINISH SELECTION, MANUFACTURING TECHNIQUES AND OTHER FACTORS AFFECTING FILTER BAG PERFORMANCE IN LIGNITE FIRED BOILER APPLICATIONS

Charles E. Capps
Menardi-Criswell Corporation
3908 Colgate
Houston, Texas

ABSTRACT

Filter bags are a small part of the overall cost of operating a fabric filter. however, they are one of the most important factors in the successful operation of the unit.

Filter bags should be designed for specific applications. Many variables must be considered. This paper will address three areas which we consider of prime importance.

First, the fabric selection. Through the years a number of fiber glass fabrics have been developed for utility boiler applications. This paper will address the fabric design and how it relates to bag life and performance.

The second important variable which must be considerd is the finish. There are in use today, three groups of finishes with many variations in each. These are the Teflon based finishes, those described as acid resistant finishes and the tri-component finishes. This paper will discuss the function, composition, application and use of these various finishes.

The final factor to be addressed will be filter bag fabrication. This paper will highlight those areas which are not normally considered by the end user when purchasing or developing a spcification for a high temprature fiber glass filter bag.

ECONOMIC ANALYSIS OF SO_2 CONTROL BY DRY SORBENT INJECTION

T. P. Dorchak

U.S. Department of Energy
Morgantown Energy Technology Center
Grand Forks Project Office
Grand Forks, North Dakota 58202

and

W. S. Lanier and S. B. Robinson
Energy and Environmental Research Corporation
200 Eastowne Drive, Suite 203
Chapel Hill, North Carolina 27514

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for
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Grand Forks Project Office
P. O. Box 880
Morgantown, West Virginia 26505

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ECONOMIC ANALYSIS OF SO2 CONTROL BY DRY SORBENT INJECTION

by

T.P. Dorchak, W.S. Lanier and S.B. Robinson

ABSTRACT

An economic analysis has been performed evaluating furnace injection of calcium based dry sorbent into lignite-fired utility boilers. A baseline analysis considered sorbent injection at an existing 500 MW_e boiler firing low sulfur North Dakota lignite. The primary objective of this baseline analysis was to evaluate the relative economics of three levels of on-site sorbent processing assuming that the sorbent being injected was a pressure-hydrated, high calcitic material. The first option examined was direct purchase of the pressure hydrate. The second option was to procure pebble quicklime and to incorporate on-site pressure hydration into the system design. The third option considered purchase of limestone with on-site calcination and pressure hydration. The baseline analyses were performed at assumed sorbent injection rates sufficient to achieve 50 to 70 percent SO_2 control while firing the low sulfur baseline coal. The lower injection rate is representative of an SO_2 reduction level which may be required by pending acid rain legislation while the higher injection rate would provide sulfur control commensurate with EPA's New Source Performance Standard. Results from this baseline analysis indicate a significant economic advantage for purchase of pebble quicklime with on-site pressure hydration. This sorbent processing option requires capital expenditure for on-site pressure hydration equipment but these costs are more than offset by the lower unit cost of lime relative to purchased pressure hydrate. The option of purchasing limestone with on-site calcination and pressure hydration becomes increasingly attractive with increasing sorbent consumption. The study indicates that lime requirements above approximately 500 tons per day are required before this option is economically viable.

Economic evaluations of the type presented in the current study are potentially sensitive to many of the inherent assumptions in the analysis. Sensitivity analyses are included evaluating the impact of assumed boiler size, boiler capacity factor, coal sulfur content, sorbent unit cost and calcium utilization efficiency. This type of analysis is of maximum value in comparative assessments.

Disclaimer

INTRODUCTION

A three-phase program has been conducted by the Energy and Environmental Research Corporation (EER) under contract DE-AC18-84FC10616 to the U. S. Department of Energy. The purpose of the program was to evaluate furnace injection of dry calcium-based sorbents to reduce SO_2 emissions from lignite-fired boilers. The first two program phases provided for field tests to determine the effects of sorbent injection on SO_2 reduction, boiler performance and electrostatic precipitator performance. The field tests were conducted at Otter Tail Power Company's Hoot Lake Unit 2 which is a 53.5 MWe tangentially-fired boiler burning North Dakota lignite. Results from the field tests are documented in earlier reports.[1,2] The third phase of the program, summarized in the current report, consisted of an economic evaluation of sorbent injection technology.

The process being evaluated involves injection of pressure-hydrated lime sorbent into the upper furnace of a lignite-fired utility boiler. Within the furnace, the injected sorbent particles dehydrate to form highly porous lime particles with high surface area, which react with SO_2 to form solid calcium sulfate. The captured sulfur is collected either in an electrostatic precipitator or baghouse and removed from the boiler exhaust. The ash along with spent and unreacted sorbent can be disposed of using normal utility practice.[3]

A series of baseline economic analyses were performed to estimate the cost of retrofitting sorbent injection to a hypothetical 500 MWe boiler firing North Dakota lignite. Results from the field test studies were used to estimate the required sorbent injection rate to achieve a desired SO_2 control level. In the baseline study, analyses were performed for retrofits designed to achieve both 50 and 70 percent SO_2 control. The fifty percent control level was selected as representative of the SO_2 reduction likely required by pending acid rain legislation. Seventy percent SO_2 control is the level required by EPA's New Source Performance Standard (NSPS) for low-sulfur coal-fired boilers. At each control level, an easy retrofit situation requiring minor plant modification and a difficult retrofit requiring ESP replacement were considered in order to bracket the range of possible costs.

A second aspect of the baseline economic analysis was an assessment of three on-site sorbent processing options. The simplest sorbent injection system design option is the direct purchase of the pressure-hydrated sorbent from a vendor. A second option is the purchase of pebble quicklime which requires on-site pressure hydration of the sorbent. The third option is the purchase of limestone and provision for on-site calcination and pressure hydration.

As described above, the baseline economic analysis consisted of a parametric evaluation examining:

- 50 and 70 percent SO_2 control,
- three levels of on-site sorbent processing, and
- easy and difficult retrofits.

This resulted in a matrix consisting of twelve cases as shown in Table 1. All costs developed for each baseline case as defined represent complete costs for capital charges, both direct and indirect, and levelized costs inclusive of tax, debt and stockholder burdens along with inflationary factors.

Table 1. Baseline Evaluation Matrix

CASE	INJECTED SORBENT	SO2 REMOVAL	RETROFIT TYPE
1 2 3	Purchased Pressure Hydrate Purchased Lime Purchased Limestone	50%	Easy
4 5 6	Purchased Pressure Hydrate Purchased Lime Purchased Limestone	70%	Easy
7 8 9	Purchased Pressure Hydrate Purchased Lime Purchased Limestone	50%	Difficult
10 11 12	Purchased Pressure Hydrate Purchased Lime Purchased Limestone	70%	Difficult

In another portion of the economic evaluation, a series of sensitivity analyses were performed examining the impact of process and boiler variables on the cost of sorbent injection SO_2 control. Included was an evaluation of the impact of boiler size, boiler capacity factor, coal sulfur content, sorbent unit cost and the calcium utilization efficiency of the injected sorbent.

ECONOMIC ANALYSIS METHODOLOGY

The current economic assessment of dry sorbent injection has adopted the economic analysis methodology defined by the Electric Power Research Institute in their Technical Assessment Guide.[4] Detailed capital cost estimates were developed including a series of contingency factors and accounting for the spectrum of construction and preproduction costs. The capital costs are presented, normalized to the boiler rating ($/kW installed capacity). Incremental operating and maintenance costs are calculated on a generalized first year operational basis and are expressed as mills per kilowatt hour. The current study focuses on technology retrofit. It is assumed that the boiler being retrofitted has a 15 year remaining book life. Economic parameters used to calculate capital and O&M cost levelizing factors are defined in Table 2. After levelization, the capital and O&M costs are combined and expressed as the incremental millage rate. The actual cost of the control technology is defined as the levelized annual cost normalized by the mass of SO_2 removed, expressed in dollars per ton SO_2 removed.

Table 2. Major Economic Premises

Book Life, Years	15
Tax Life, Years	5
Inflation Rate, %/Year	6.0
Debt/Equity Ratio	50/50
Debt Cost, %/Year	11.0
Preferred Stock Ratio, %	15
Preferred Stock Cost, %/Year	11.5
Common Stock Ratio, %	35
Common Stock Cost, %/Year	15.3
Weighted Cost of Capital, %/Year	12.5
Federal Plus State Income Tax Rate, %	50
Property Taxes and Insurance, %/Year	2
Investment Tax Credit, %	10
Levelized Carrying Charge Rate, %	16.01
O&M Cost Levelizing Factor	1.4517

BASELINE BOILER ANALYSIS

A 500 MWe boiler firing a North Dakota lignite was selected as the baseline boiler for evaluating the cost of retrofitting dry sorbent injection. The assumed boiler and lignite characteristics are defined in Table 3. Based on these parameters, the uncontrolled SO_2 emission rate is calculated as 5.10 tons/hour (at 100% load), or 29.07 ktons/year.

The baseline evaluation considered the economic impact of three system variables on the cost of dry sorbent injection SO_2 control. Included are the impacts of (1) the extent of on-site sorbent processing, (2) the percentage SO_2 control, and (3) the ease of technology retrofit. The study is based on use of pressure-hydrated, high-calcitic lime as the dry sorbent material injected into the boiler. The three sorbent processing approaches described previously were used in designing the technology retrofit.

As part of the design study, various calciner design options were also considered and an annular shaft kiln was selected for the sorbent injection technology economic analyses.

The second major system variable considered in the baseline analysis is the percentage SO_2 control achieved by sorbent injection. Control levels of 50 and 70 percent were selected. The data presented in Figure 1 illustrate the variation in percent SO_2 capture as a function of sorbent injection rate (expressed as Ca/S molar ratio) as taken from field tests at Hoot Lake. In these tests, high-calcitic, pressure-hydrated sorbent was injected into a utility boiler burning the lignite defined in Table 3. Based on this field test data, 50 percent SO_2 capture is achieved at calcium-to-sulfur molar ratio of 2.0, while 70 percent SO_2 capture requires calcium-to-sulfur ratio of 4.2. Thus, to increase the SO_2 capture from 50 to 70 percent requires a 110 percent increase in sorbent injection rate and a commensurate increase in size of sorbent storage and handling hardware.

The major system parameter considered in the baseline evaluation was the ease with which the required retrofit could be accomplished. There are many

Table 3. Baseline Boiler and Fuel Characteristics

BOILER CHARACTERISTICS

Size	500 MW_e
Net Plant Heat Rate	10,600 Btu/kWh
Capacity Factor	65%
Bottom Ash-Fly Ash Split	20%-80%
Excess Air-Air Heater Inlet	20%
-ESP Inlet	35%
ESP	Sized to Meet 1979 NSPS for Particulate on Baseline Coal.
ESP Inlet Temperature	300°F

FUEL ANALYSIS

PROXIMATE	AS RECEIVED	DRY
Moisture	35.0%	—
Ash	7.0%	10.8%
Volatile Matter	26.7%	41.1%
Fixed Carbon	31.3%	48.1%
HIGHER HEATING VALUE (Btu/lb)	6,890	10,600
ULTIMATE ANALYSIS		
Moisture	35.0%	—
Carbon	42.9%	66.0%
Hydrogen	2.85%	4.4%
Nitrogen	0.6%	0.9%
Sulfur	0.78%	1.2%
Ash	7.0%	10.8%
Oxygen	10.87%	16.7%

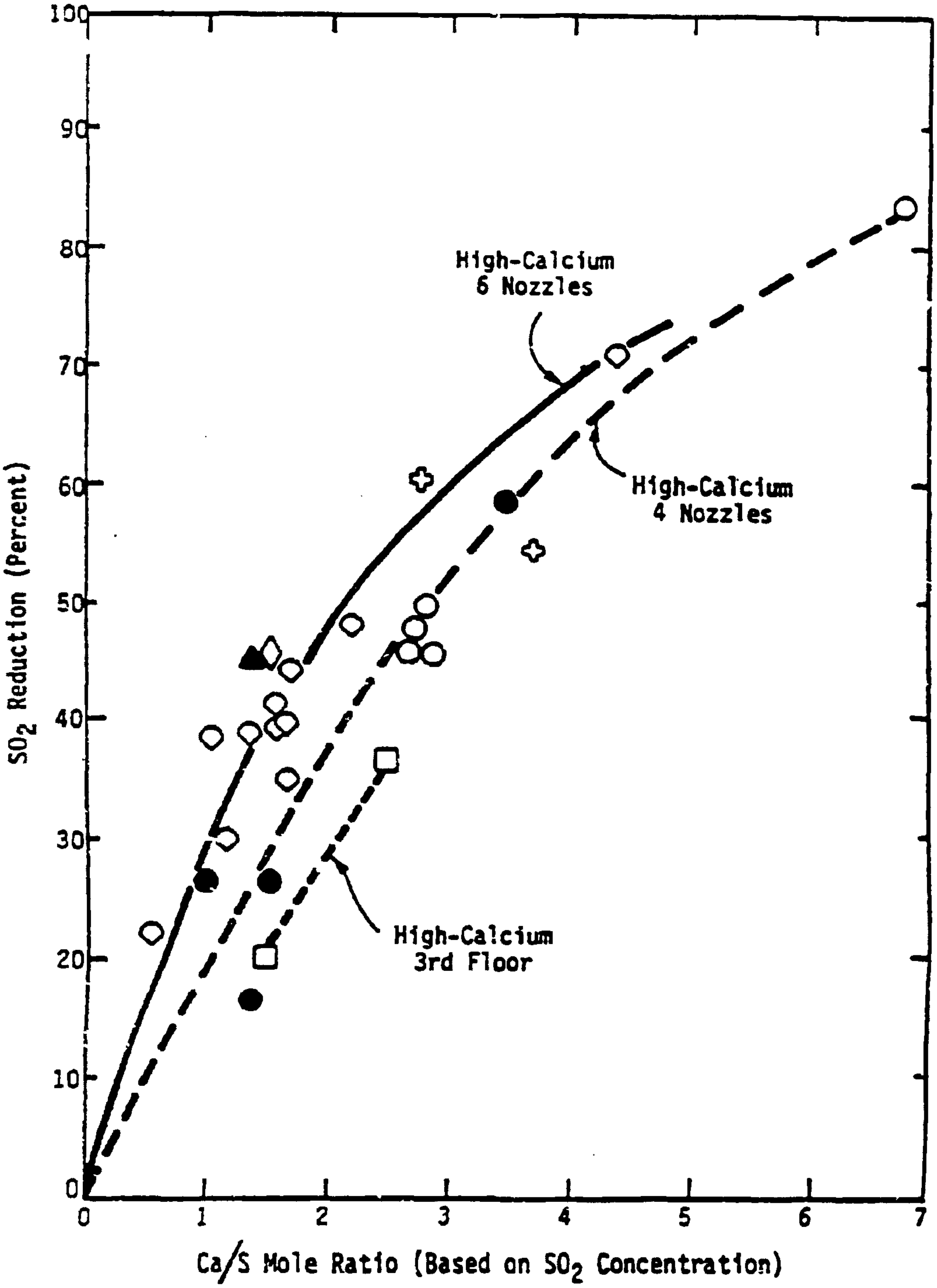

Figure 1. Summary of SO_2 Reduction Data from Field Test on Otter Tail Power's Hoot Lake Unit No. 2, from Ref. 1

boiler designs, facility layouts, and fuel characteristics which will impact retrofit cost for dry sorbent injection. One of the most significant factors, however, is whether the boiler's existing particulate control device will be capable of handling the increased particulate loading and potentially increased resistivity due to SO_3 scavenging by the sorbent. In the previously noted field tests on Hoot Lake Unit Number 2, satisfactory ESP performance was maintained under all operating conditions. This particular boiler/coal combination would represent an "easy" retrofit since no boiler or APCD modifications were required. It is possible, however, that situations could exist requiring complete replacement of the existing particulate control hardware and upgrading of soot blowing equipment. This would clearly represent a "difficult" retrofit situation. The current study attempts to determine the potential range of technology costs by defining an "easy" retrofit as a situation requiring no upgrade to the soot blowing cycles or the the APCD hardware. Conversely, a "difficult" retrofit would require an increase in the number of soot blowers as well as replacement of the existing ESP with a baghouse. The increased pressure drop of a baghouse relative to an ESP would also require additional fan capacity. For the difficult case, capital cost estimates were increased by a factor of 1.3 times that of an equal sized new unit installation.

Sorbent Handling/Processing - The hardware configuration assumed for sorbent unloading, long-term storage, processing and transport to the boiler for each of the three sorbent processing options is illustrated in Figure 2. Figure 2a illustrates the layout for direct purchase of pressure hydrate and shows the hardware for delivery to the site by railroad car, long-term storage in covered, concrete silos, day bins and pneumatic transport to the boiler. The number and size of the long-term storage silos is adjusted to reflect provision for a 30-day on-site sorbent supply. Figure 2b illustrates the hardware requirements for on-site pressure hydration while Figure 2c illustrates on-site calcination and pressure hydration. These general layouts are directly patterned after the hardware requirements defined by the Tennessee Valley Authority in their economic assessment of lime and limestone based flue gas desulfurization.[5] In fact, where applicable, the capital cost for these system components, including equipment and installation, were directly extracted from TVA's Shawnee economic analysis model. Capital costs for components not covered by the TVA model were determined by vendor quotes or other standard estimation procedures. In each evaluation case sufficient redundancy has been assumed, commensurate with utility industry practice.

Boiler Modifications - For easy retrofit cases, the only boiler modifications required for sorbent injection retrofit are (1) movement of minor interferences to get sorbent transfer lines to the boiler face, (2) a series of wall penetrations to allow sorbent injection and (3) addition of a control system to meter and modulate process operation. Movement of interferences is estimated to cost approximately \$0.3/kW installed capacity while an appropriate control system is estimated to cost approximately \$300,000. For the baseline boiler, it is assumed that twelve (12) wall penetrations will be required at a cost of approximately \$10,000 per penetration. For a difficult retrofit there will be costs associated with increased soot blowing capability. A recent EPA economic evaluation[6] of the sorbent injection process estimated the capital cost of additional soot blowers at \$1.09 per kilowatt of installed capacity. That assumption has been adopted in the current study as well.

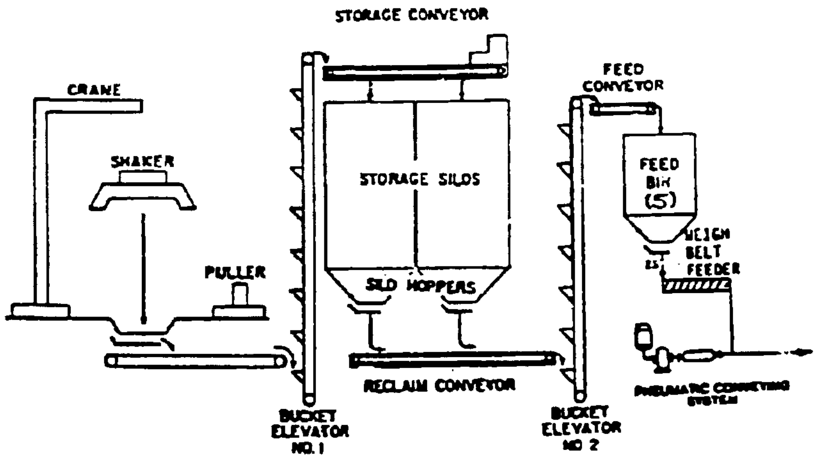

Figure 2a. Equipment Layout for Direct Purchase of Pressure Hydrate

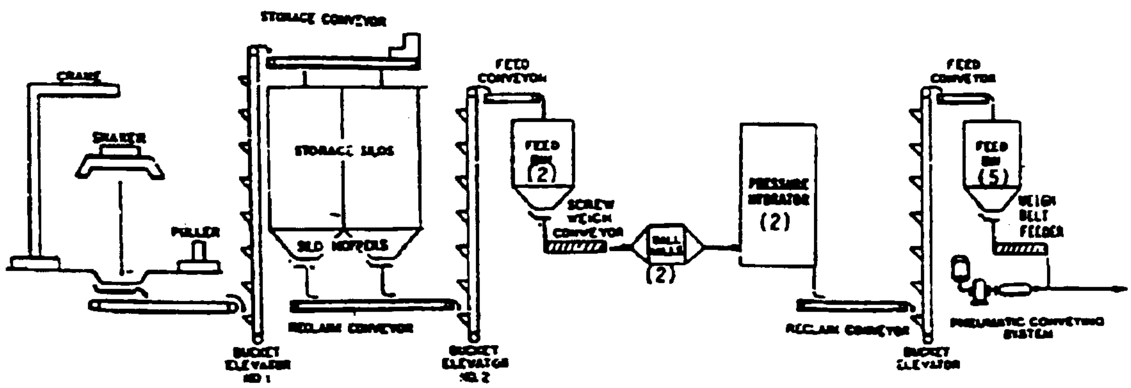

Figure 2b. Equipment Layout for On-site Pressure Hydration

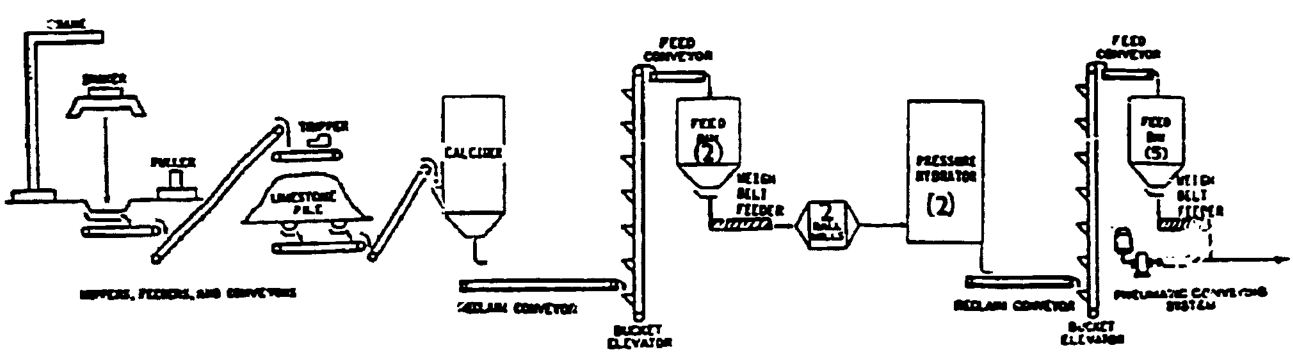

Figure 2c. Equipment Layout for On-site Calcination and Pressure Hydration

O&M Cost Estimation Bases - As will be shown, a critical component of the overall sorbent injection control technology cost is the O&M component. Annual consumption quantities for sorbent, electricity, water, incremental solids disposal, operating labor and analytical labor are calculated (or estimated) for each individual evaluation case and multiplied by unit costs to estimate direct operating costs. Also included in the direct operating cost are charges for added hardware and maintenance which is estimated as four percent of the direct construction cost. Indirect operating costs are taken as 60 percent of maintenance labor and materials, operating labor and analysis labor. The sum of direct and indirect operating costs represents the generalized first year O&M cost. Unit costs assumed for the baseline analysis are defined in Table 4. As will be shown, sorbent injection technology economics are dominated by the cost of consumed sorbent and thus the unit cost for sorbent becomes a critical evaluation parameter.

Table 4. Unit Costs for Baseline Analysis

Parameter	Unit Cost
In-House Electricity	4.5 cents/kWh
Water	50 cents/kgal
Natural Gas	$3.75/10^6 Btu
Operating Labor	$18.50/hour
Analysis Labor	$25.00/hour
Baseline Fly Ash Disposal	$4.25/ton
Fly Ash + Spent Sorbent Disposal	$6.25/ton
Pressure-Hydrated, High-Calcitic Sorbent	$80.00/ton
Pebble Quick Lime	$55.00/ton
Limestone	$20.00/ton

Easy Retrofit Economic Assessments - Table 5 presents a summary of the capital costs for baseline cases 1,2 and 3 as defined in Table 3. As shown, the capital costs range from $16.77/kW for direct purchase of pressure hydrate to $36.56/kW for on-site calcination and pressure hydration. Table 6 defines the O&M costs for cases 1,2 and 3 as well as presenting the levelized cost summary. As shown, for 50 percent SO_2 capture with an easy retrofit, the lowest retrofit cost is achieved by on-site pressure hydration of purchased quicklime. It is significant to note that direct purchase of pressure hydrate provides the lowest capital cost but that this option provides the highest overall cost as indicated by cost per ton of SO_2 removed and levelized revenue requirements.

The same general trends are observed for cases examining easy retrofit designed for 70 percent SO_2 capture. Specific information noted is that on-site calcination is still more expensive than purchase of pebble quicklime but that the cost penalty is decreased. The general trend noted here is that on-site calcination becomes economically attractive only when the lime use rate exceeds approximately 500 tons per day. This finding is in general agreement with information provided by lime manufacturers.

Difficult Retrofit Cases - As indicated previously for the difficult retrofit cases it is assumed that the existing ESP will be replaced with a baghouse. For pricing purposes it is assumed that the replacement baghouse provides an

Table 5. Capital Cost Summary for Case 1,2 and 3:
Easy Retrofit, 50 Percent SO2 Control

DIRECT CONSTRUCTION COSTS (DCC)	Case 1 ($1,000's)	Case 2 ($1,000's)	Case 3 ($1,000's)
1. Sorbent Unloading, Storage and Reclaim	3,384	1,923	1,259
2. Calcination and Lime Storage	—	—	4,074
3. Pressure Hydration and Sorbent Storage	—	3,778	4,283
4. Sorbent Injection	281	281	281
5. Wall Penetrations	120	120	120
6. Movement of Interferences	150	150	150
7. Control System	300	300	300
	4,235	6,551	10,466
Process Contingency	423	655	1,047
DCC TOTAL	4,658	7,206	11,513
INDIRECT CONSTRUCTION COST (ICC)			
General Facilities	466	721	1,151
Engineering and Home Office	466	721	1,151
Project Contingency	1,397	2,162	3,454
ICC TOTAL	2,329	3,604	5,756
TOTAL PLANT COST (DCC + ICC)	6,987	10,810	17,269
ALLOWANCE FOR FUNDS DURING CONSTRUCTION	0	0	0
TOTAL PLANT INVESTMENT	6,987	10,810	17,269
INITIAL INVENTORY	689	358	233
PREPRODUCTION	711	625	779
TOTAL RETROFIT CAPITAL COST	8,387	11,793	18,281
	(16.77/kW)	($23.58/kW)	($36.56/kW)

Table 6. O&M Cost Summary for Case 1,2 and 3:
Easy Retrofit, 50 Percent SO_2 Control

	Case 1		Case 2		Case 3	
	Quantity	Cost ($1000s)	Quantity	Cost ($1000s)	Quantity	Cost ($1000s)
ITEM						
Sorbent Consumption	67.2kTPY	5372	50.82	2795	90.75	1815
Utilities						
Electricity	876,000kWh	39	2,917,000	131	4,187,000	188
Natural Gas	—	—	—	—	177,866	667
Water	—	—	5,616kgal	3	5,616	3
Incremental Solids Disposal	—	677	—	677	—	677
Maintenance Labor & Material						
Operating Labor	13,000hr/yr	240	27,000	499	35,000	647
Analysis Labor	1,000hr/yr	25	1,000	25	2,000	50
Direct Operating Cost		6,541		4,419		4,508
Indirect Operating Cost		271		487		695
1st Year O&M Cost		$6,812		$4,906		$5,203
1st Year Revenue Requirement		2.88mills/kWh		2.39mills/kWh		2.86mills/kWh
15 Year Levelized Revenue Req.		3.96mills/kWh		3.16mills/kWh		3.68mills/kWh
Control Technology Cost		$755/ton		$620/ton		$721/ton

air-to-cloth ratio of 2.0 $ACFM/ft^2$. Algorithms for projecting the capital cost of baghouses have been developed by PEI Associates, Inc.[7] and indicate that a new unit sized for the baseline boiler (including booster fans) would cost approximately $17 million. After applying a 1.3 retrofit factor it is estimated that the incremental direct construction cost for a difficult retrofit would be approximately $22.6 million. Addition of various contingencies and other factors estimated as a fraction of direct construction cost indicate that the incremental total retrofit capital cost is approximately $37.4 million or $74.91/kW. These are increments to capital costs calculated for an easy retrofit.

As regards O&M costs, the simplifying assumption was made that the cost of operating and maintaining a baghouse will be approximately the same as that for an ESP. On this basis, the difference between easy and difficult retrofit will be reflected only in the capital cost increment.

Base Case Summary - Table 7 provides an overall cost summary for all twelve base case evaluations indicating capital costs, O&M costs, 15-year levelized revenue requirements, and control technology costs. As indicated in this table, the most attractive option is to purchase quicklime and to perform on-site hydration. The genesis of that advantage lies mainly in the fact that lime maximizes the calcium content in the delivered material. This in turn minimizes shipping costs. If the lime is hydrated at the manufacturer's facility, transportation costs are incurred for a significant weight of water.

The economic evaluation of the twelve baseline cases also illustrated that the cost of sorbent injection will depend heavily on the ability of the existing particulate collection hardware to maintain acceptable performance. Additional research is required to define appropriate modifications in the event that significant ESP performance deterioration does occur. It should be noted that several organizations, including the Department of Energy, are investigating duct humidification as a potential modification and preliminary results are reportedly encouraging in terms of increased sorbent capacity, sulfur capture and ESP performance.

SENSITIVITY ANALYSIS

Units of 250 MW_e and 750 MW_e were examined to determine the sensitivity to boiler size, assuming that the baseline lignite is used as fuel and that the utility is purchasing lime with on-site pressure hydration. For a 250 MW_e boiler, the lime consumption rate is reduced to 4.47 tons per hour. This reduced rate allows for unloading of the rail cars with a pneumatic system, and a single concrete storage silo is sufficient for storing a 30-day supply of lime. However, due to lack of turndown capabilities on the pressure hydrators, the equipment requirements for pressure hydration would remain the same as for a 500 MW_e installation. The retrofit cost for the situation described above is estimated to be $42.4/kW, and the control technology cost is estimated to be $806 per ton of SO_2 removed.

For a 750 MW_e boiler retrofit, the lime consumption rate is increased to 13.39 tons per hour. The lime unloading system will be the same as the 500 MW_e system, though three concrete silos will be required for a 30-day storage capacity. The full-load sorbent injection rate of 17.69 tons per hour requires three pressure hydrators, although the sorbent storage and injection

TABLE 7. BASE CASE SUMMARY

CASE NO.	PURCHASED REAGENT	RETROFIT TYPE	SO_2 CAPTURE percent	CAPITAL COST $/kW	O&M COST mills/kWh	15 YR LEV. REV. REQMT. mills/kWh	CONTROL TECH. COST $/ton SO_2
1	$Ca(OH)_2$	EASY	50	16.77	3.47	3.96	755
2	CaO	EASY	50	23.58	2.50	3.16	620
3	$CaCO_3$	EASY	50	36.56	2.65	3.68	721
4	$Ca(OH)_2$	EASY	70	27.87	6.69	7.47	1046
5	CaO	EASY	70	36.98	4.42	5.46	764
6	$CaCO_3$	EASY	70	56.64	4.52	6.11	855
7	$Ca(OH)_2$	DIFFICULT	50	91.56	3.47	5.06	1167
8	CaO	DIFFICULT	50	98.37	2.50	5.26	1032
9	$CaCO_3$	DIFFICULT	50	111.35	2.65	5.78	1133
10	$Ca(OH)_2$	DIFFICULT	70	102.66	6.69	9.57	1458
11	CaO	DIFFICULT	70	111.77	4.42	7.56	1176
12	$CaCO_3$	DIFFICULT	70	131.43	4.52	8.21	1267

hardware used for the 500 MW_e system will be sufficient for the 750 MW_e system. An economic analysis of this system indicates a capital cost of \$23.6/kW, and a control technology cost of \$594 per ton of SO_2 removed.

The sensitivity of cost to variations in boiler capacity factor was also examined. The baseline boiler was assumed to operate at 65 percent capacity factor. This is typical of a base-loaded unit, though many units are intermittently loaded and operate at significantly lower capacity factors. All boilers must be designed to meet full-load operation, however, regardless of capacity factor. Therefore, the capital cost of retrofit for SO_2 control will not vary with this parameter. Most of the O&M costs do vary linearly with this parameter, and thus the influence of capacity factor on sorbent injection control technology cost will depend on the relative contribution of these two factors.

Figure 3 illustrates the influence of capacity factor on 15-year levelized revenue requirement and control technology cost for a 500 MW_e boiler. Figure 4 portrays the same information for a 250 MW_e boiler. The cost of sorbent injection is seen to increase with reduction in capacity factor for both boiler sizes, but the increase is greater for the 250 MW_e unit. This is primarily due to the higher relative capital cost for the smaller unit. It should be noted, however, that SO_2 control technologies with high relative capital costs, such as FGD, will be even more adversely affected by low capacity factors.

The costs for sorbent in the baseline analysis were assumed to be \$80/ton for pressure hydrate, \$55/ton for quicklime, and \$20/ton for limestone. This study examined the effect of an incremental change in sorbent cost on control technology cost. Figure 5 presents these results for an easy retrofit to the baseline boiler at 50 percent SO_2 capture. This figure shows the control technology cost for lime purchase with on-site hydration to be \$628 per ton of SO_2 removed. To achieve an equivalent control technology cost with direct purchase of pressure hydrate would require a hydrate cost reduction to \$64.30 per ton. Using limestone with on-site calcination and hydration would require a limestone unit cost of approximately \$9.33 per ton.

Finally, an economic assessment was conducted to determine the sensitivity of sorbent injection control technology costs to calcium utilization rates ranging from 15 to 40 percent. Figure 6 illustrates the impact of this variation for 50 percent SO_2 control. Here the Ca/S ratio is not held constant, but is dictated by a given calcium utilization rate. This figure shows that significant improvement in sorbent injection economics may be realized through the development of more reactive sorbents, although increased sorbent costs may partially offset the reduction in control technology cost.

CONCLUSION

The results of this study indicate a significant economic advantage for the purchase of lime with on-site pressure hydration. This sorbent processing option requires capital expenditure for on-site pressure hydration equipment, but these costs are compensated for by the lower unit cost of lime as opposed to purchased pressure hydrate. This economic advantage holds for both easy and difficult retrofits, and for systems designed for either 50 or 70 percent SO_2 removal. The purchase of limestone with on-site calcination and pressure

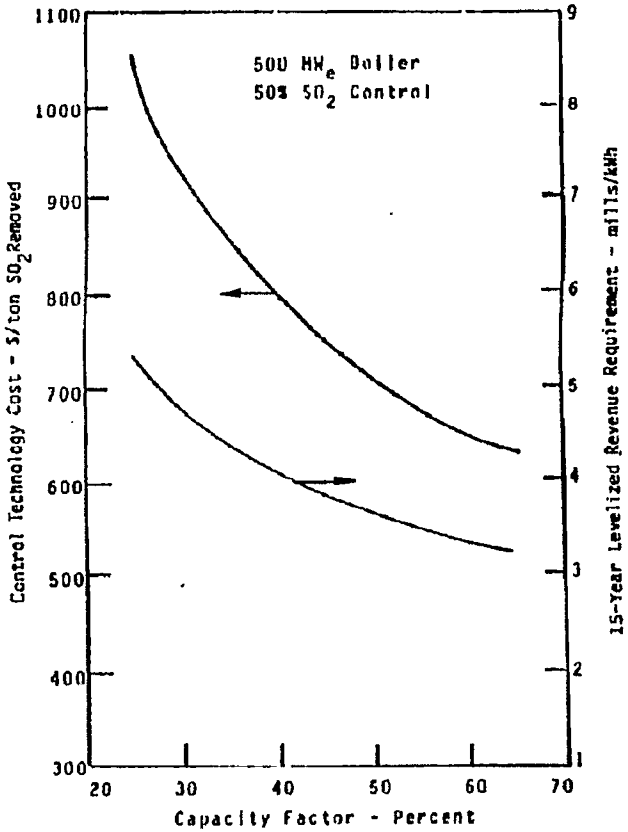

Figure 3. Influence of Capacity Factor on the Cost of Sorbent Injection. Easy Retrofit to 500 MW$_e$ Boiler

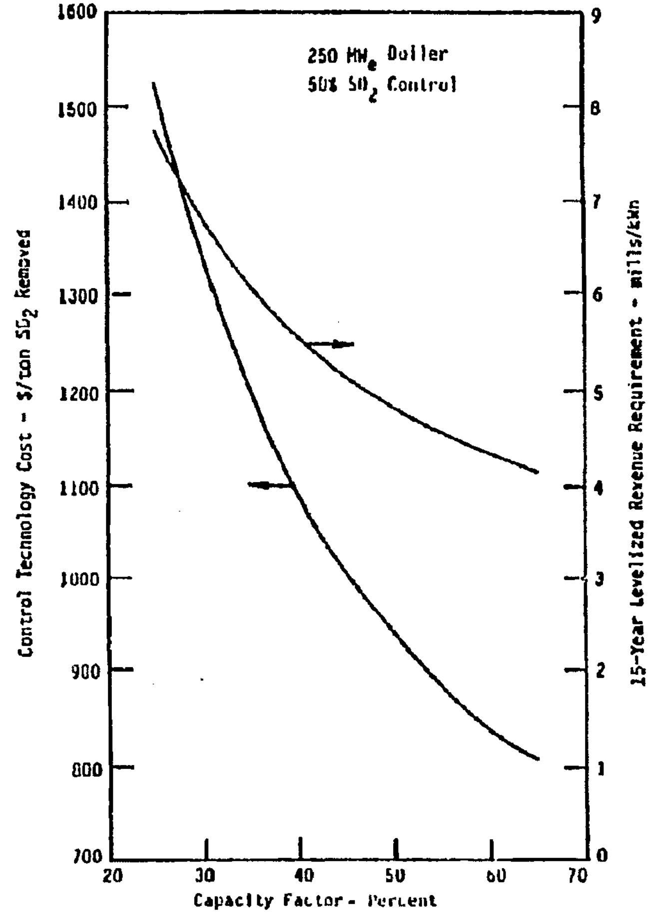

Figure 4. Influence of Capacity Factor on the Cost of Sorbent Injection. Easy Retrofit to a 250 MW$_e$ Boiler

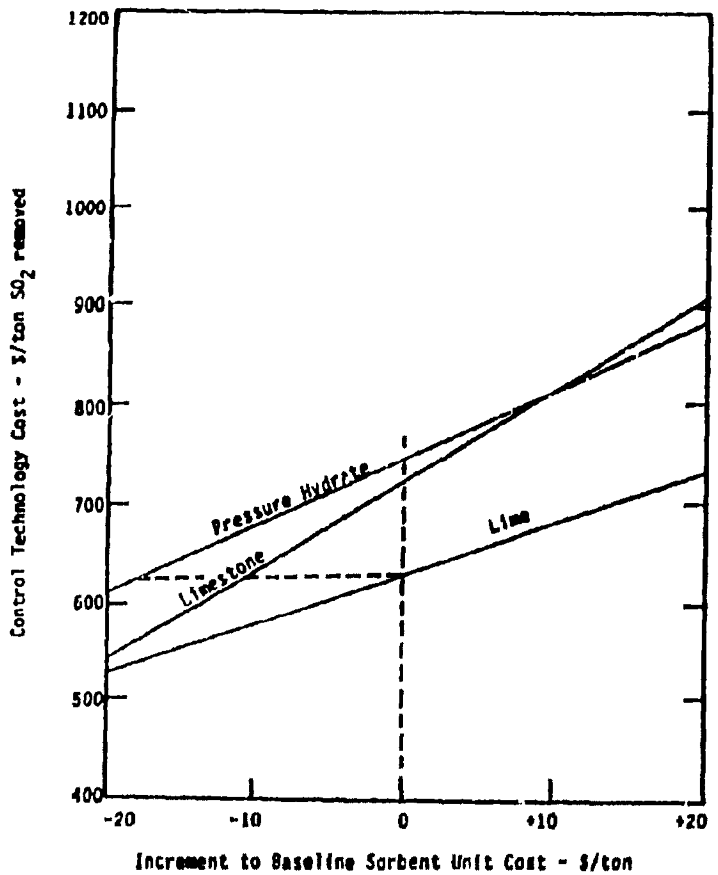

Figure 5. *Influence of sorbent unit cost on sorbent injection control technology cost. Easy retrofit to 500 MW_e with 50 percent SO_2 capture.*

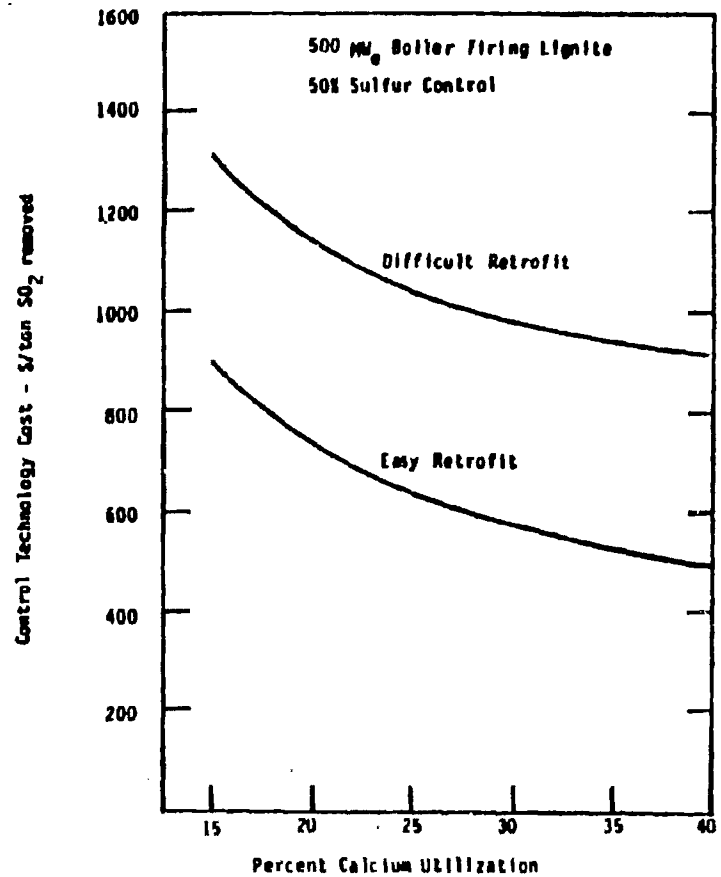

Figure 6. Influence of Calcium Utilization rate on sorbent injection control technology cost.

hydration is economically attractive only for retrofits requiring more than 500 tons per day of lime. This will occur only for large boilers which fire relatively high-sulfur coals.

The sensitivity of sorbent injection costs to a variety of parameters was examined. Boiler size above 500 MW_e has relatively minor influence on the economics. A 250 MW_e boiler has approximately 30% higher control technology costs than a 500 MW_e boiler, though units over 500 MW_e show little decrease. The capacity factor has a large impact on the control technology cost, and the magnitude of the impact is dependent on the relative contribution of capital and fixed operating costs to the total revenue requirement. Calcium utilization rates, particularly below 25%, can have dramatically negative impact on the economics. A corollary to this is the very significant sensitivity shown to sorbent unit cost. Overall sorbent injection technology is particularly attractive for boilers firing low-sulfur coals. Other comparative economic studies[8] have also resulted in this conclusion.

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OPTIMIZATION OF THE GIBBONS CREEK LIMESTONE FGD SYSTEM

Prepared by:

J. D. Colley
R. D. Delleney
A. F. Jones

Radian Corporation
Austin, Texas

Prepared for Presentation at:

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ABSTRACT

The Texas Municipal Power Agency operates a 440 MW, mine-mouth lignite boiler in East Central Texas. This unit, Gibbons Creek 1, is equipped with a cold-side ESP and a limestone spray tower FGD system to control emissions from the combustion of a nominal 1% sulfur lignite. The FGD system experienced numerous operating and reliability problems until TMPA initiated an optimization program in late 1984. As a consequence of TMPA's characterization of the FGD system with the assistance of Radian Corporation and the subsequent implementation of many of the recommendations that resulted from this characterization, significant reductions in operating and maintenance costs have been achieved. The system operability and reliability have also been improved. This paper presents the results of the characterization study and a discussion of the FGD system improvements that have been realized.

1.0 INTRODUCTION

The Gibbons Creek 1 Station of the Texas Municipal Power Agency (TMPA) is a 440 MW, mine-mouth lignite unit. The plant is located in east central Texas near Bryan. Unit 1 fires a 1.06% sulfur lignite and is equipped with an emission control system comprised of a cold-side ESP, followed by three limestone spray towers. Operation of Unit 1 began in 1982.

Radian Corporation completed a characterization of the Texas Municipal Power Agency (TMPA) flue gas desulfurization system at Gibbons Creek during late 1984. The scope of the characterization included both a chemical and engineering evaluation of the performance and design of the FGD system. Liquid, slurry, and solid phase samples were collected around the scrubbers, and process and design data on the FGD system were collected. The primary objective of the characterization was to obtain the necessary information to develop an approach for improving the reliability and reducing the operating and maintenance costs of the Gibbons Creek scrubbers. Based on the results of the characterization, a list of recommendations to improve the performance of the scrubbers was developed and is in the process of being implemented by TMPA. Recent FGD performance is reviewed in the final section.

2.0 CHARACTERIZATION TEST RESULTS

2.1 SO_2 Removal

The Gibbons Creek FGD system was designed to control SO_2 emissions with two operating towers and a third for spare under normal lignite sulfur levels. Table 1 shows the number of towers and spray pumps required at varying sulfur loadings to the FGD system. According to analyses of the lignite burned at Gibbons Creek, the average sulfur content was 1% in the fuel, with a range of 0.8% to 1.2%. Comparing this to the CE design, two-tower operation with a total of six or seven spray pumps in service should have controlled SO_2 emissions to below the 1.2 lbs SO_2 per million Btu standard. Because of various problems, including higher than expected tower pressure drops, mist eliminator scaling and pluggage, and plugged spray nozzles, three towers and a total of nine spray pumps were operated the majority of the time.

In addition, the pH setpoint in the reaction tanks was maintained above 6.0, since below this value the tower removal efficiencies reportedly dropped off rapidly. At the time of the Radian characterization test, the pH was not controlled closely and the average value measured was about 6.5. The FGD system was reportedly designed to operate at a limestone stoichiometric ratio of 1.1. As discussed in more detail in Section 2.2, this ratio was measured to be above 1.5.

The FGD system, as currently operated, is capable of maintaining the unit in compliance with the SO_2 standard with the current fuel quality. There is significant room for improvement in terms of reducing the scrubber operating and maintenance costs, however. Specifically, there are three areas where significant cost reductions are possible:

TABLE 1. EFFECT OF SULFUR LEVEL ON DESIGN SPRAY PUMP AND TOWER REQUIREMENTS - GIBBONS CREEK

Fuel Sulfur		CE Design		Total Treated Gas Flow[1] (acfm)	Total Recycle Flow (gpm)	Design L/G[2]
lbs SO_2/ 10^6 Btu	% Sulfur	Number of Towers	Number of Pumps			
4.4	0.9	2	6	1,050,000[3]	109,000	104
9.5	2.0	2	10	1,360,000[4]	182,000	134
11.8	2.5	3	15	1,360,000[5]	273,000	200

[1]Actual conditions - 142°F, saturated.

[2]gals/10^3 acf.

[3]Assumed tower SO_2 removal efficiency of 92%; therefore, 23% bypass.

[4]Assumed tower SO_2 removal efficiency of 88%; therefore, no bypass.

[5]Assumed tower SO_2 removal efficiency of 92%; therefore, no bypass.

1. Limestone utilization;
2. Number of operating spray pumps; and
3. Scaling and pluggage in the mist eliminators.

With several modifications, operation of the unit with two towers and six spray pumps, good limestone utilization (greater than 90%), and reduced mist eliminator scaling should be possible. The modifications include:

1. pH feedback control of limestone to the scrubbers;

2. Operation at a constant pH of 5.8 to 6.0 to maintain a limestone stoichiometric ratio of 1.1;

3. Upgrading the ball mill circuit so that it produces a product of at least 90% less than 325 mesh;

4. Washing the mist eliminators with fresh water;

5. Installation of density control instrumentation for the reaction tanks; and

6. Installation of a screen to remove sticks from the limestone slurry.

An additional consideration in achieving the minimum operating and maintenance costs for the FGD system is examination of the feasibility of using dibasic acid (DBA) in the scrubbers. A preliminary evaluation indicates that adding DBA would:

1. Allow two-pump operation per tower;

2. Improve reliability of the mist eliminators by reducing secondary scrubbing of SO_2 in the mist eliminators;

3. Allow 95% utilization of limestone in the scrubbers; and

4. Maximize the amount of bypass for reheat, thereby, minimizing the amount of steam needed to maintain the desired stack temperature.

A preliminary cost/benefit analysis was completed for the use of DBA in the scrubbers. The analysis was based on the operation of the scrubbers as observed during the characterization testing (three towers, three pumps per tower, 64% limestone utilization). Table 2 presents the results of the cost/benefit analysis. A number of assumptions had to be made to complete the analysis. Table 3 presents these assumptions.

TABLE 2. INITIAL COST/BENEFIT ANALYSIS FOR GIBBONS CREEK FGD SYSTEM OPTIMIZATION

DBA	Annual Cost
Feed System Amortization	$ 42,000
Raw Material	425,000
Subtotal	$ 467,000
Benefit	**Annual Benefit**
Limestone	$ 730,000
Limestone Milling	40,000
Sludge Disposal	240,000
Pump Power	774,000
Fan Power	120,000
Reheat	100,000
Mist Eliminator Maintenance	100,000
Pump and Nozzle Maintenance	50,000
Subtotal	$2,154,000
NET ANNUAL BENEFIT	$1,687,000

TABLE 3. BASIS FOR COST/BENEFIT ANALYSIS

Maximum Unit Load	400 MW
Annual Capacity Factor	0.90
Busbar Cost of Electricity	\$0.045/KW-hr
Reduction in Number of Operating Spray Pumps	6
Previous Limestone Stoichiometric Ratio	1.54
Limestone Stoichiometric Ratio with DBA	1.05
Cost of Limestone	\$13/ton (delivered)
Power for Grinding Limestone	14 KW-hr/ton
DBA Consumption Rate	30 lbs/ton SO_2 absorbed
Cost of DBA Feed System (The cost of the DBA feed system is amortized over 15 years at 12% interest.)	\$150,000
Cost of DBA	\$0.18/lb (delivered)
Maintenance Costs Due to Mist Eliminator Scale	\$100,000/yr

The results of the analysis showed a substantial cost reduction is possible. Much of the 2.1 million dollar total benefit can be realized without the use of DBA. For example, controlling the pH alone at 5.8-6.0 in the reaction tanks has resulted in an annual savings of slightly over 1 million dollars by reducing limestone consumption, disposal, and milling costs. The improvement in limestone utilization should also benefit mist eliminator reliability. The primary advantage that DBA offers is in reducing the number of operating spray pumps. The 600-hp spray pumps use 3.5×10^6 KW-hrs of electricity annually, and the 700-hp spray pumps use 4.1×10^6 KW-hrs per year. Assuming TMPA could sell this power at $0.045/KW-hr, the 600-hp pumps cost $160,000/year and the 700-hp pumps cost $185,000/year to operate, not including maintenance costs. By reducing the number of operating spray pumps from the current level of six down to four with DBA addition, the net station power output would be increased by 1 megawatt. A proportional reduction in nozzle and pump maintenance would also result. The annual delivered cost of DBA is estimated to be about $467,000, including the amortized cost for the feed system based on the San Miguel test results. The benefit to TMPA of adding DBA would be about $900,000 in increased power sales and a reduction in spray pump and nozzle wear. The net benefit associated with DBA after deducting the cost for the raw material would be over $500,000 annually.

2.2 Limestone Utilization

Proper control of limestone utilization in a limestone scrubber is very important for two reasons. First, the lower the utilization is, the greater the amount of limestone sent to disposal and wasted. Second, Radian experience has seen the effect of utilization on mist eliminator scaling and pluggage. Maintaining a utilization consistently above 90% is important in helping to prevent scaling and pluggage in the mist eliminators. As will be discussed later, it is not the only ingredient to good mist eliminator performance, though.

The results of the slurry analyses indicated that limestone utilization was generally very low. Analysis of the filter cake, which represented a good composite of FGD system performance over a several day period, showed that limestone utilization was only 64%. This converts to a stoichiometric ratio of 1.56. This number was used for the calculations presented in the previous section. The average of the separate reaction tank samples was even lower than this, at just over 60%. The reason for the low utilization was due to the high reaction tank pHs, which were due in turn to the lack of control of limestone fed to the tanks. Figure 1 is a plot showing the relationship between limestone utilization and pH based on slurry data. The general shape of the curve is the same as Radian has observed at other utility limestone FGD systems.

The key to maintaining a stable concentration of limestone in the reaction tanks is to control the pH in the tanks at a constant value. Use of the pH to control limestone addition to the reaction tanks in a feedback mode of operation was recommended. Upgrading the current pH measurement system to improve the accuracy and reliability of the pH signal was also needed.

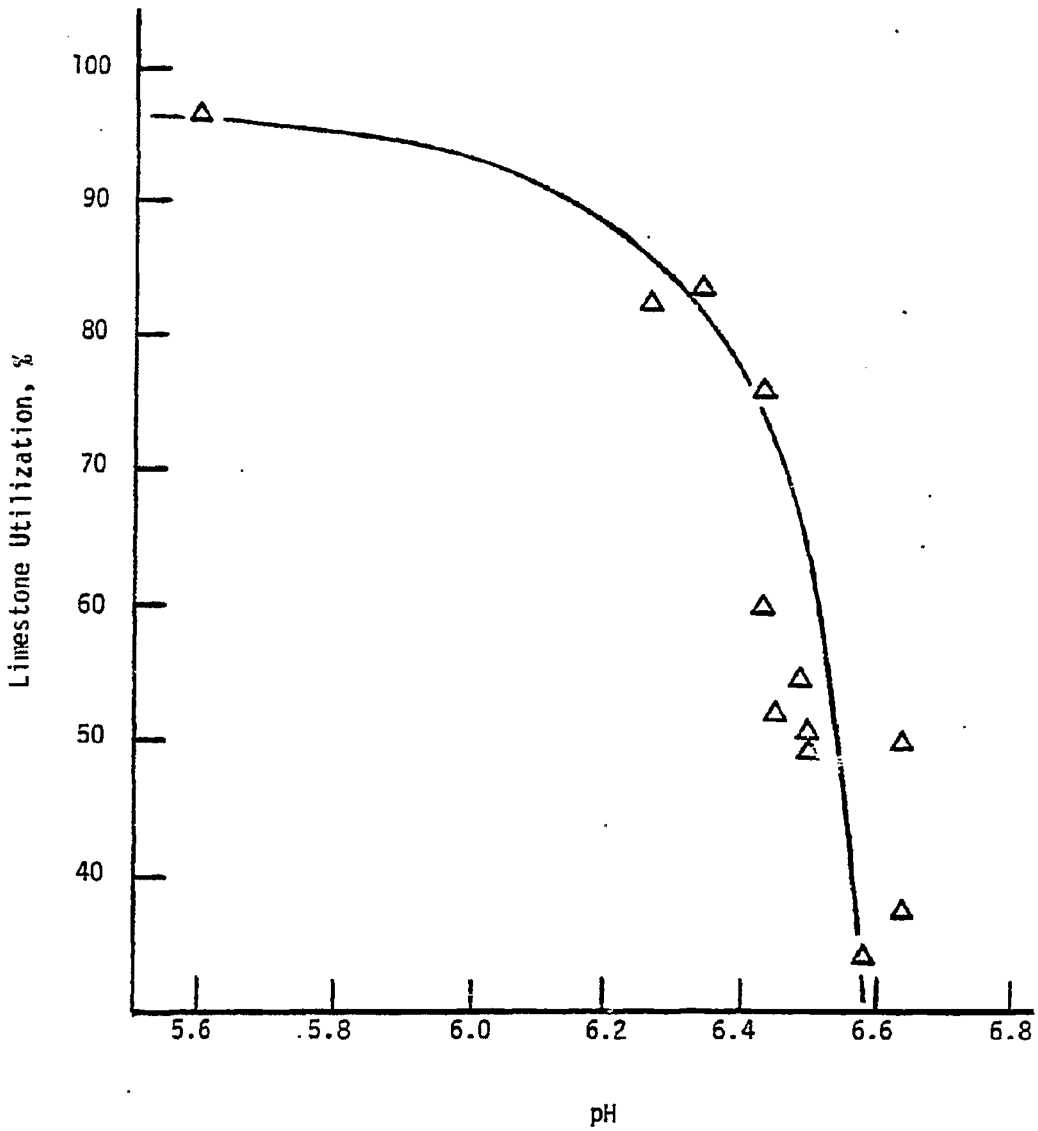

Figure 1. Effect of Reaction Tank pH on Limestone Utilization

In the recommended pH feed back control loop, the signal from the pH transmitter is used to control the operation of the limestone make-up value in an on/off fashion. A controller receives the 4-20 ma output from the pH transmitter, and then maintains the reaction tank pH within a deadband of 0.2 pH units by opening the make-up valve at low pH (5.8) and closing it at high pH (6.0). This has worked successfully at a number of limestone FGD systems that Radian has worked with.

Further improvement in scrubber performance would also result from upgrading the ball mill circuit to produce a finer limestone product. Measurement of the product size showed that 83% passed a 325 mesh screen. This categorizes the product as a medium size. Based on observations of the current operation and design of the mill circuit and conversations with the cyclone vendor (Krebs), it is possible that the system may be easily modified to produce a product of 92% to 95% passing a 325 mesh screen. Past research with spray towers has seen an improvement in SO_2 removal by feeding the system a finer limestone. Radian has documented the effect of finer limestone utilization in scrubbers. Experience at Duck Creek (Central Illinois Light Company) has shown that the finer the limestone is, the higher the allowable pH setpoint to maintain a given utilization. For example, by feeding the Gibbons Creek scrubbers a finer limestone, the reaction tank setpoint may be moved up to 6.0-6.2 and still achieve better than 90% utilization. The advantage would be that the scrubbers could operate at higher SO_2 removal efficiencies at the higher pH setpoint. In addition, the finer limestone is less erosive, and therefore, pipe and nozzle life may be extended.

2.3 Mist Eliminator Scaling

Scaling and plugging in the mist eliminators needed to be eliminated. In addition, mist carryover from the partially plugged mist eliminators might have contributed to the corrosion problems in the reheaters. Clean mist eliminator operation should reduce carryover rates. As a result of the characterization testing, Radian identified two potential causes for the scale formation in the mist eliminators. The first was poor limestone utilization in the scrubbers. The second potential cause was the use of thickener overflow and/or ash pond water to wash the mist eliminators.

The major cations and anions were analyzed in the liquid samples collected around the FGD system. The analyses were input to the Radian Liquid Equilibrium Program to predict the calcium sulfate (gypsum) relative saturation. Relative saturation is a convenient means for evaluating the scaling potential of a particular liquor or slurry. A guideline that has been proposed for acceptable washwater quality is no greater than 0.5 gypsum relative saturation. Both the ash pond and the thickener overflow had a relative saturation higher than this. Furthermore, the best success with keeping mist eliminators clean has been seen with the use of a fresh washwater source. In addition, the quantity of mist eliminator wash required can be kept to a minimum if good quality washwater is used.

2.4 FGD System Water Balance

A significant overflow from the thickener to the ground was observed during the characterization period. Hydraulic constraints were reportedly responsible for this since there was not sufficient surge capacity within the system to even out large influxes of slurry from the scrubbers and filtrate from dewatering to the thickener.

Several changes which could alleviate or perhaps solve this problem were identified. First, the control of the density in the reaction tanks should be put into automatic operation using a density instrument and controller to regulate blowdown from the scrubbers. The setpoints for the controller should be selected so that large volumes of slurry are not discharged to the thickener. This can be done by narrowing the deadband on the controller so that it maintains density within a fairly narrow range of specific gravity (1.060 to 1.065). In addition, the hydraulic loading on the thickener could be further reduced by installation of controls to prevent two or three scrubbers from discharging to the thickener at the same time.

Second, the level instruments could be used to automatically control the make-up water rate to the scrubbers. In this way, as the scrubbers discharge slurry to the thickeners on density control, the resulting drop of the level in the reaction tanks will trigger the make-up valves to open and return displaced liquor from the thickener to the reaction tank.

In addition, increasing limestone utilization should significantly improve thickener operation, as well as the hydraulic balance. The reason can be seen in Table 4. Table 4 shows the effect of operating with improved utilization on the volumetric rate of blowdown from the scrubbers. By operating with less limestone fed to the scrubbers at higher utilizations, the blowdown rate to the thickener is substantially reduced.

TABLE 4. EFFECT OF LIMESTONE UTILIZATION AND SLURRY DENSITY ON SCRUBBER BLOWDOWN RATES

Limestone Utilization (%)	Reaction Tank Density (Wt. %)	Slurry to Thickener (gpm)	Solids in Blowdown (tons/hr)
64	10	1030	27
91	10	780	21
95	15	490	20

The blowdown to the scrubbers could be cut by more than half in comparison to the previous operation. The result would be a noticeable improvement not only in the hydraulic balance, since much less water and slurry would be moving from the scrubbers to the thickener, but also a significant increase in solids settling time in the thickener. Polymer dosage rates to the thickener could possibly be reduced as a result.

If DBA is used in the Gibbons Creek scrubbers, it will be very important to operate the FGD system with an overall negative water balance. Any water that is discharged from the FGD system would carry dissolved DBA with it, increasing the DBA feed rate to the scrubbers. For example, at a DBA concentration of 1000 ppm in the scrubber slurry, a 50 gpm continuous discharge of water from the FGD system would increase the annual cost for DBA by $40,000. Table 5 presents a rough estimate of the Gibbons Creek overall FGD water balance.

The best available information indicates that the current FGD water balance is negative by 80 gpm. However, if the source of the mist eliminator washwater is switched to fresh water to help keep it clean, the second case shows that a positive water balance would result. As pointed out earlier, if DBA was being added to the scrubbers, this would result in a significant loss of the additive. The third case presents what the balance would look like if the source of limestone grinding was switched to thickener overflow and the mist eliminators were washed with fresh water. In this case, the water balance is negative by almost 120 gpm. Case IV shows the effect of converting all packing gland pump seals to mechanical seals. Radian is involved in several programs sponsored by EPRI or private utilities comparing the technical and economic advantages and disadvantages of several different types of mechanical seals to packing seals on large volume absorber feed pumps.

3.0 FGD SYSTEM IMPROVEMENTS

In 1985, TMPA began implementation of several of the recommendations outlined in Section 2.0. The most significant were the institution of a different control scheme for feeding limestone to the scrubbers and adhering to a specific pH setpoint for the reaction tanks. The objective of these changes was to improve the limestone utilization in the FGD system and to improve system reliability and reduce operating and maintenance costs.

In March 1985, Radian again visited the Gibbons Creek Station and reviewed the installation and operation of the new limestone control system. Samples of the reaction tank slurry were collected and analyzed to quantitatively measure the level of limestone utilization. The results of this test indicated that the scrubbers were averaging about 90% utilization at a pH setpoint of 5.8. The scrubbers were achieving the goal that had been set for them. Furthermore, the limestone control system appeared to be operating in a very stable and accurate mode. This was due in part to the new state-of-the-art pH measuring equipment and in part to the novel approach

TABLE 5. GIBBONS CREEK FGD SYSTEM WATER BALANCE

	Water In (gpm)			Water Out (gpm)		
Case Description	Limestone Grinding	Pump Seal Water	Mist Eliminator Wash	Vaporized	With Sludge	Net Balance (gpm)
I. Current Operation	190	140	50[1]	400	60	-80
II. Current Operation with Fresh Water Mist Eliminator Wash	190	140	200	400	60	+70
III. Fresh Water Mist Eliminator Wash with Recycle Water for Limestone Grinding	---	140	200	400	60	-120
IV. Fresh Water Mist Eliminator Wash with Recycle Water for Limestone Grinding and Mechanical Pump Seals	---	0	200	400	60	-260

that TMPA was using to interface the pH electrodes with the process slurry. This novel approach involved mounting the pH probe in the side of the reaction tank through a valve, making maintenance and calibration easier.

An added benefit of improving the limestone utilization was a simultaneous reduction in the rate of scaling in the scrubber mist eliminators. In the past, Radian has seen a strong relationship between limestone utilization and mist eliminator scaling. Although the scaling was not totally eliminated as hoped for, further work is continuing to stop all scaling in the tower mist eliminators by converting the mist eliminator wash to a fresh water source.

Additional improvements that have been or are being implemented by TMPA include the use of self-limiting orifices in the seal water feed system, the use of thickener overflow (recirculated water) for limestone grinding, and the use of reclaim water for wash down. These changes result in an optimized use of fresh water in the system for areas such as the mist eliminator wash while maintaining the system water balance to avoid FGD system discharges. Some of the numerous other areas where work is ongoing include the nozzle size and design, the use of modulating valves in the limestone feed lines, an evaluation of new mist eliminator designs, the use of bypass reheat, absorber and outlet duct materials, and the installation of gravity flow bleed lines from the secondary reaction tanks to the thickener.

One final area where TMPA and Radian are working together is the use of DBA as an FGD additive. As mentioned in Section 2.0, DBA would improve the liquid phase alkalinity of the scrubber liquor. As a result of increasing the alkalinity of the liquor, the SO_2 removal of the scrubbers will increase. If sufficient amounts of DBA are present then the removal efficiency will be raised enough so that a spray pump can be turned off and the unit remain safely in compliance. Increasing the DBA concentration even more will further improve alkalinity to the point that more spray levels can be turned off.

As spray levels are turned off, the power consumption of the station decreases and the power may be sold instead of consumed by the operating spray pumps. With fewer pumps on-line, the spray nozzle, motor, and pump maintenance will decrease in proportion to the number of spray levels turned off. In addition, the reduction in spray levels will lower the overall scrubber pressure drop and thereby reduce the boiler ID fans' power consumption. Furthermore, with fewer pumps on-line, the amount of mist entrained in the scrubbers and carried up to the mist eliminators is reduced. This should help prevent scaling in the mist eliminators. Since the initial optimization in 1984, TMPA has been able to reduce the number of spray pumps in operation from a total of nine down to six. This has been possible through improved limestone addition control and the feeding of a finer limestone to the towers. Also, TMPA has made modifications to reduce the number of plugged nozzles, thereby improving gas/liquid contacting.

Based on the current baseline operation of the scrubbers, which incorporates the improved control of the system, a DBA cost/benefit analysis was completed to assess whether this additive could further optimize the TMPA scrubbers. The results are included in Table 6. This analysis assumes that with DBA only two scrubbers and two pumps per tower would be needed for compliance operation. It also assumes that sufficient flue gas bypass is possible to allow elimination of the in-line reheat system. For example, if the scrubbers were controlled at an SO_2 removal efficiency of 95% (they currently operate at about 84%) through the addition of DBA, approximately 18% of the flue gas could be bypassed around the scrubbers for reheat. This would supply about 30 degrees of reheat, allowing the stack to operate at about 175°F. The assumptions used for the pump power and reheat savings were updated to reflect the current TMPA cost credits for power and steam. These are $17/MW-hr for pump power and $1.48/1000 lbs of steam.

TABLE 6. DBA COST/BENEFIT ANALYSIS FOR TMPA GIBBONS CREEK

DBA	Annual Cost
Feed System Amortization	$ 40,000
Raw Material	210,000
Subtotal	$ 450,000
Benefit	**Annual Benefit**
Pump Power	400,000
Reheat Steam	290,000
Nozzle Maintenance	15,000
Pump and Motor Maintenance	15,000
Subtotal	$ 720,000
NET ANNUAL BENEFIT	$ 270,000

The result indicates that the use of DBA may be beneficial for further optimization of the scrubbers. However, before the major credit for reheat can be taken, the carbon steel ductwork from just past the reheat coils to the stack must be lined to prevent corrosion. TMPA is currently examining lining the ductwork independently due to corrosion of the carbon steel even with the reheat system in operation. Another consideration is that operation without direct or indirect reheat capabilities means that with higher sulfur fuel (which is seen periodically), the amount of bypass possible will decrease to perhaps only 10% and the stack temperature would drop below 170°F.

Determination of the exact level of benefit from the addition of DBA to the scrubbers can only be made by testing. The actual DBA feed rate required to achieve two-tower, two pump per tower operation with 18% bypass must be measured over at least a four- to six-week period. The consumption data and annual DBA cost can then be compared to the observed benefits to determine the attractiveness of using this additive on a long-term basis.

SO_2 SCRUBBER PERFORMANCE - "THE BOTTOM LINE"

by Mike Wadlington, Manager of Technical Services-Chemistry
and Bryan Ferguson, Chemical Specialist - Scrubbers
TU Electric Generating Division

TU Electric Generating Division operates 12 lignite fired steam generating units at four locations: Monticello, Martin Lake, Big Brown and Sandow Steam Electric Stations. Five of these units (three at Martin Lake, and one each at Monticello and Sandow) have SO_2 scrubber systems treating the flue gas. The first of these scrubber systems was placed in service on Martin Lake Unit No. 1 in 1977. The last was placed in service on Sandow unit No. 4 in 1981. These scrubbers, sometimes referred to as Flue Gas Desulfurization (FGD) systems, are all of the limestone slurry-spray tower type but have some significant differences as designed by the three separate vendors that provided the original equipment. Original Equipment Manufacturers (OEM's) were Research-Cottrell at Martin Lake, Chemico at Monticello, and Combustion Engineering at Sandow.

All three systems are described, including major process/hardware changes and subsequent improvements. Operating experiences are summarized especially with regard to scrubber performance monitoring. Performance indices were developed and are utilized to evaluate SO_2 compliance and Megawatt limitations due to the scrubbers. Operating and maintenance costs for the past three years are also compared.

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FGD SYSTEM DESCRIPTIONS

Martin Lake. The Martin Lake Steam Electric Station is located approximately 130 miles southeast of Dallas near the town of Tatum. Units No. 1, No. 2, and No. 3 were placed in operation in April 1977, May 1978, and February 1979, respectively. Each unit is equipped with a Research-Cottrell Double Loop FGD System of eight towers (originally six). Two additional towers per unit and forced oxidation were retrofitted in 1983 to increase the gas and solids treating capacity so that higher sulfur lignite could be burned in the boilers.

Figure 1 (Top) shows the general arrangement of the absorber towers. Each tower pair, or module, is fed by a common absorber feed tank (AFT). Flue gas at a flow rate of 3.9 million ACFM and 360°F enters the inlet manifold to the scrubber towers. Based on inlet sulfur levels, from 75 to 95 percent of the flue gas is treated while the remaining 5 to 25 percent of the flue gas by-passes the system to the outlet duct. A schematic of a typical absorber tower is shown in Figure 1 (Bottom). Flue gas enters the quencher where it is contacted by slurry and cooled to a temperature of about 140°F.

Quenched flue gas then passes through the liquid gas separator commonly referred to as the "bowl" into the absorber section. The bowl separates the absorber loop slurry from that of the quencher loop. In the absorber section, the flue gas passes through absorber sprays, wetted film contactor (Primary Packing), and a set of mist eliminators (ME). The mist eliminators remove entrained moisture from the treated flue gas prior to its entering the outlet duct where it mixes with any by-passed flue gas and exits to the stack.

As designed and initially operated, the mist eliminator wash frequency was controlled to maintain a constant slurry density in the AFT's. Limestone slurry was added to maintain the pH setpoint and all slurry overflowing the AFT went to the quenchers. The density in the quenchers was controlled by discharging slurry to the thickeners on high density. The level in the quencher tanks was controlled by returning thickener overflow to the tank.

Monticello. The Monticello Steam Electric Station is located approximately 110 miles east of Dallas near Mt. Pleasant. Units No. 1 and No. 2 were placed in service December 1974 and December 1975, respectively. No FGD systems were required for these units. Monticello Unit No. 3 was started in August 1978 and is equipped with three open spray towers provided by Chemico Air Pollution Control Corporation (now General Electric Environmental Services, Inc.).

Figure 2 (Top) shows the general arrangement of the FGD spray towers and a flow diagram of an absorber is shown in Figure 2 (Bottom). After passage through the electrostatic precipitator for particulate removal, three centrifugal boiler I.D. fans drive the flue gas at 360°F and 3.6 million ACFM, at full load, into a common inlet manifold from which it can be equally distributed into the spray towers for SO_2 removal. The scrubbed gas leaving the spray towers is again collected in a common manifold together with any by-passed gas prior to entering the stack.

The SO_2 scrubbing is accomplished by three self-supporting spray towers with integral slurry recycle tanks. The gas enters the spray towers from the bottom inlet, takes a 90° turn and moves upward through four sets of counter-current sprays followed by chevron type mist eliminators for removal of liquid entrainment. The four banks of spray nozzles produce a large amount of reagent spray droplets thereby creating the effective surface area for mass transfer of the SO_2 from the gas. The drops then fall into the reaction tank (488,000 gallons capacity per tank) from which it is recycled. Limestone make-up slurry is added to the reservoir (reaction tank) at the bottom of the tower to maintain proper pH for SO_2 removal. Slurry is kept in suspension by four agitators which also aid in maintaining reaction equilibrium. Absorber recycle slurry is bled under pressure from the recycle pumps to maintain density control and is disposed of in an on-site pond. The reclaimed water from the disposal pond is recycled to the FGD system.

Top and bottom wash sprays are provided to prevent solids deposition on the mist eliminators. The mist eliminator is divided into twelve pie shaped segments. Two opposing sections are sprayed sequentially on an adjustable timed cycle. The mist eliminators are washed either with service water or recycle water.

Sandow. The Sandow Steam Electric Station is located about 60 miles east of Austin near Rockdale, Texas. Units No. 1-3 were constructed in the early 1950's by Alcoa to generate electricity for an adjacent aluminum plant and TU Electric was contracted as operator of the plant. Sandow Unit No. 4 is owned by TU Electric and was placed in operation in May 1981. The unit is equipped with three open spray towers provided by Combustion Engineering (C-E). Space was provided for a fourth tower if increased sulfur levels of lignite were to be utilized.

A schematic of a typical absorber tower is shown in Figure 3 (Bottom). The C-E Scrubber utilizes both ladder type vanes within the module and a perforated plate at the absorber inlet to direct the 2.5 million ACFM flue gas. These devices are intended to aid in straightening, proportioning, and directing the gas flow upwards into an effective velocity profile. Sandow Unit No. 4 has four spray levels per absorber that discharge the slurry counter current to the gas flow creating a blanket of atomized droplets at each level through which the gas passes. Retention time in the absorber is provided by the towers effective height of 31 feet which when constructed was the largest on any C-E FGD system. As the flue gas leaves the spray portion of the absorber with entrained water droplets, it passes through the demister section which is composed of a bulk entrainment separator (BES) followed by a two-stage chevron mist eliminator. The treated flue gas exits from the absorber into the outlet duct where it mixes with any by-passed flue gas. An external ambient air reheat system was also supplied to allow 100% gas scrubbing capability. The BES and chevron vanes are washed with recycled scrubber pond water by four washer lances located between the bulk entrainment separator and the lower mist eliminator. Six venturi nozzles set at a 45 degree angle in the washer lance, coupled with a 360 degree lance rotation are designed to provide contact of each vane by a water jet.

The absorbers are top supported so that the bottom of the absorber shell (the chute) hangs in its own reaction tank. In addition to providing a liquid holding area conducive to the completion of the chemical reaction and providing slurry to the recycle pumps, forced oxidation occurs in the reaction tanks (437,000 gallons capacity per tank) using air spargers to complete conversion of calcium sulfite to calcium sulfate. Primary process control was based on specific gravity of the scrubber liquor with limestone slurry being added to the reaction tank and waste slurry pumped to a settling pond.

FGD OPERATING EXPERIENCE

Martin Lake. The ability to pass boiler gases through the scrubber towers dictated to a large extent, the ability to operate the Martin Lake Units in compliance with SO_2 emission standards at maximum output. Gas flow through a scrubber tower can vary due to the pluggage of that towers packing. By far, the most costly problem with Martin Lake's scrubber towers in terms of both maintenance expense and loss of generation due to compliance backdown was tower packing pluggage. The primary packing, first stage demister, and the second stage demister all saw pluggage due to the loss of flow through associated spray nozzles and the scaling of gypsum due to excursions in the process chemistry.

Prior to late 1981, the major identified cause of pluggage in primary packing was the entrainment of solids as the nozzles over the packing plugged with debris and scale picked up by the pumps feeding the nozzle headers. This pluggage problem was addressed by the installation of pump suction screens in late 1981. The screens, tested and recommended by the Scrubber O.E.M., greatly reduced nozzle pluggage which doubled the scrubber system's capacity and availability to treat flue gas. The problem that still remained to be addressed was the scaling and pluggage due to chemical imbalance.

TU Electric developed a maintenance program which involved manual cleaning of each tower at approximately 40 to 60 day intervals. A tower would be taken off-line and isolated using its dampers, and a crew of men would then remove plugged and damaged packing. The mist eliminators often required cleaning and replacement since they also were scaled and plugged. With this maintenance frequency, begun in late 1981, the scrubber system could be operated so that few load reductions were necessary to meet SO_2 compliance. The expense associated with the more concentrated maintenance program, however was very large.

In September 1982, TU Electric formed a task force to evaluate, prioritize and recommend courses of action to improve the operation of the scrubber system. This task force was composed of representatives from the plant, the corporate engineering and technical support groups, the FGD system's O.E.M., and independent consultants. TU Electric was invited by EPRI to participate in the FGD Process Troubleshooting Program in mid-November 1982. A brief chemical and process characterization of the FGD system was performed by Radian, EPRI's contractor, in March 1983. After review by the FGD Task Force, a test program was recommended to demonstrate the effectiveness of proposed equipment and operational changes to reduce the scale formation in the towers. The test program was conducted from December 1983 through April 1984. Based on the results of the EPRI test program, it was found that a split limestone feed/gypsum recycle modification would minimize chemical scale formation in the FGD system. The modification consisted of a separate feed line for limestone slurry to the quenchers and a controller to regulate addition of limestone to the AFT and quenchers together with a recycle of a portion of slurry from the quencher recycle loop to the AFT module. See Figure 1 (Bottom). The gypsum recycle increased the gypsum crystal concentration in the AFT to provide a greater number of crystallization sites for the calcium sulfate formed in the towers. With its feedback control loop, the split limestone feed system controlled the addition of limestone to both the AFT's and quenchers to maintain a constant pH set point in both.

The decision was made to install the split limestone/gypsum recycle system on all three units during the Fall 1984 outages. The new system included new pH transmitters, piping and control valves for the modification, and a more durable primary packing developed by Poly Manufacturing. The tighter control of process operation that resulted coupled with an extensive analytical program enhanced both operation and performance allowing the production of a commercial grade of gypsum to be produced and marketed as a by-product.

Monticello. In the initial months of operation, several breaks in the fiberglass line that supplies reclaim water from the sludge disposal pond to the towers were experienced due to inadequate support and restraint of piping. This had been the only problem that resulted in the removal of the scrubber when the generator was on-line and the problem was corrected with the replacement of the fiberglass line with carbon steel pipe.

The more significant problem experienced with the FGD system was the repeated failures of the rubber lining of the slurry recycle pumps. Although the problem did not result in the loss of availability of the FGD system or noncompliance with emission limits, it became necessary to operate with all three towers in service when part of the recycle pumps were out of service. The recycle pumps that experienced failure were 800 hp pumps rated at 16,000 GPM. Both the impellers and linings experienced massive damage. After extensive tests and trials with various pumps and linings the substitution of polyurethane as a lining material has been shown to give the best results for this system's requirements.

Another problem experienced has been the use of ash water and pond water for mist eliminator wash. The high levels of calcium sulfate resulted in extreme fouling of the mist eliminator packing material even while using various scale inhibitors. The high velocity of gas through the unplugged areas combined with the increased load on other towers results in slurry carryover into the outlet duct and chimney. In addition, the use of blended water often encouraged overwashing with little improvement and some detrimental effects to density control and plant water management. To improve the reliability of the mist eliminators, the wash system is being upgraded by replacing the wash nozzles and eliminating leaks in the wash system valving. The nozzle arrangement is also being modified to improve distribution, so that areas that receive little coverage will now be adequately washed.

The leaking valves and resulting volume in the ME wash greatly contributed to the FGD system as a net generator of water and an excessive volume of scrubber blowdown. Rather than being able to maintain the 8-10 percent solids density, the unit averaged only 6 percent in 1986 and a blowdown rate of 700 GPM with the unit at full load. This led to problems in the return water piping from the scrubber pond due to scaling. The sulfite leaving the scrubber was oxidized to sulfate in the final pond after leaving the initial settling basins where the solids and unreacted limestone particles settle out before entering the larger scrubber pond. The resultant scrubber pond water had a gypsum relative saturation of 1.4. In general, until the relative saturation is above 1.3 supersaturation is not great enough for rapid scale formation to be seen but conversely clear liquid precipitation of calcium sulfate crystals in the pond would not be expected unless the relative saturation exceeded 2.0. The resulting tendency is for calcium sulfate to precipitate on the pipe through which the pond water is returned to the scrubbers.

The liquid sulfite concentration in the blowdown is not affected by the solids concentration. Whether the slurry density is 2 percent solids or 20 percent solids, the sulfite concentration will be constant at about 400 ppm. By returning the scrubber control to the higher density of 10 to 12 percent, less slurry and therefore less sulfite will be sent to the ponds and most of the oxidation will occur in the settling basin where sufficient particle sites for precipitation are present. The improvement in density control resulting from improved mist eliminator wash and a phase in of mechanical seals will be further supplemented by dilution with make-up water when the FGD system is no longer operating as a net generator of water.

Sandow. The Sandow Unit No. 4 FGD system was originally designed to limit SO_2 emissions to comply with the 1971 New Source Performance Standards. The design operating conditions for three different lignite sulfur levels were specified: 1.2, 1.6, and 3.0 percent. Based on the original design, two tower operation should have been sufficient to keep the unit in compliance under current fuel conditions. However, due to higher than expected gas flow rates from the boiler, all three towers are needed to maintain compliance.

During initial operation, the original rubber impellars of the recycle pumps failed to perform as desired and contributed to the pluggage of nozzle spray headers. These were eventually changed to a high chrome material. In addition, the spray header clamps continued to crack, break, and plug spray nozzles. This contributed to localized distribution problems causing localized liquid/gas (L/G), ratio problems. Even though overall tower L/G ratios were normal, localized problems can materially affect tower SO_2 removal. The problems with the spray header clamps and nozzle pluggage were greatly reduced working with the Scrubber O.E.M. to facilitate material changeouts and improvement of screens in the reaction tank.

Scaling and pluggage of the FGD system BES and ME had been experienced from intial operation and continued beyond the previous mentioned modifications due in part to the high amount of gas treated by the scrubbers and therefore the velocity of gas through the spray section and the mist eliminators. It is important not to exceed certain velocities in the spray section because of mist entrainment and mist loading on the mist eliminators. The maximum expected velocity of Sandow is about 12 ft./sec. at full load and no bypass with no pluggage. The chevron type mist eliminators app-oach their design limitations on liquid drainage at velocities in the 10 to 15 ft./sec. range. (The optimum performance in terms of mist eliminators is considered the 8 to 10 ft./sec. range.) The result of the high velocity, which increases with any pluggage, is that re-entrainment of mist often occurs on the backside of the mist eliminators with subsequent corrosion and sludge accumulation in downstream ductwork.

C-E and some Sandow personnel have also attributed the ME problems to structural problems with the mist eliminator supports rather than scaling. The supports have apparently shifted as the tower aged. The supports are now far enough apart that ME sections could fall if they completely shift to one side of the supports. Material was welded to sections of the supports to reduce the distance but there has not been sufficient time to assess this modification.

In July 1982, dibasic acid (DBA) was first used as an additive to the scrubber system where it was shown to be an effective mass transfer additive for wet limestone FGD systems. See Figure 3 (Top). Both SO_2 removal and limestone utilization were increased and allowed increased capacity without construction of an additional tower. Intermittent DBA feed has continued on an "as needed" basis. In 1985, Radian Corporation was hired to conduct an optimization program and assist in development of an approach for improving the reliability and reducing the operating and maintenance costs of the Unit No. 4 FGD system. As part of the study, it was indicated that substantial savings in spray pump horsepower could be achieved from continuous addition of DBA. A significant reduction in annual limestone cost would also result with this addition by being able to control the scrubbers at a lower reaction tank pH while still maintaining SO_2 removal.

In the data gathered by Radian, the DBA concentration to maintain compliance was found to be about 1000 ppm with a pH setpoint of 6.0 and 2 spray pumps per tower. The study also indicated the DBA consumption was approximately 74 lb. DBA per ton SO_2 removed. This is very high when compared to the 6 to 30 lb./ton of SO_2 removed range reported by other utilities. The primary reason the DBA consumption rate is high at Sandow is because the system essentially operates in an open loop mode with respect to DBA. The DBA leaves the scrubber with the disposed sludge and ends up in the scrubber pond which is so large that over five years would be needed to reach a steady state value. Therefore to make long-term continuous DBA use economically attractive, the solids concentration of the slurry discharged from the scrubbers had to be increased significantly to reduce the large amount of water and dissolved DBA which would be discharged to the ponds. The necessary increase in slurry solids concentration could not be achieved solely by controlling the reaction tanks at a higher density. Since hydroclones have been successfully applied in other FGD installations to increase the solids concentration of slurry from forced-oxidized scrubbers, a hydroclone system was designed for the Sandow scrubbers to achieve maximum recovery of DBA with a minimum of capital expenditure. This hydroclone system is being installed during 1987.

SCRUBBER PERFORMANCE

Having described our three different types of FGD systems and relating some experiences, especially with regard to improving scrubber operation, I would like to discuss the main topic of this paper "Scrubber Performance".

There are many important parameters used as indicators of FGD system performance. Among these are SO_2 removal efficiency, limestone stoichimetry, limestone utilization, FGD system availability, FGD system reliability, etc. These are all important; but I would like to suggest three parameters that are really the essential or "bottom line" indicators for evaluating Scrubber Performance. These are:

* SO_2 Compliance
* Megawatt Limitation
* O&M Cost

The only reason we have scrubbers is to meet an air quality compliance requirement; in our case 1.2 lbs. SO_2/mm BTU's, so we utilize a compliance factor defined as follows:

$$\text{Compliance Factor} = \frac{\text{Hours in Compliance}}{\text{Hours Generator On-line}} \times 100$$

Our compliance factors have improved each year as shown in the attached bar chart. (See Figure 4.) The compliance factors for all five scrubbers are over 99% for 1986.

The second major indicator we use is a Megawatt limitation factor:

$$\text{Megawatt Limitation Factor} = \frac{\text{EFOH + FOH}}{\text{EFOH + FOH + Hours Gen. On-Line}} \times 100$$

Where EFOH (Equivalent Unit Outage Hours) is load curtailment due to the scrubbers and FOH (Full Unit Outage Hours) is a unit shutdown due to the scrubbers. Spare tower redundancy at Martin Lake (eight towers per unit) as compared to Monticello Unit No. 3 and Sandow Unit No. 4 (three towers per unit) is one of the reasons for a better limitation factor at Martin Lake. The trend continues to improve with all FGD systems less than .73% for 1986.

The last performance factor is scrubber O&M cost. Substantial improvements have been made since we began monitoring O&M cost in 1984. The attached bar chart, Figure 6, shows our average O&M cost per unit of energy for all five scrubbers. The downward trend reflects reduced costs of 4.2 million dollars in 1985 when compared to 1984 and 1.8 million dollars in 1986 when compared to 1985. Our scrubber O&M cost also compares favorably with the 1.3 mils/Kwh total O&M average cost for 26 Limestone FGD Systems reported in EPRI's Report No. CS-2915.

The "bottom line" for performance monitoring of SO_2 scrubber consists of these three parameters: Compliance, Megawatt limitation, and O&M cost. TU Electric's five operating scrubbers show significant improvements in all of these areas as shown by the comparative bar charts for the last few years. The Table (Figure 7) provides a historical summary of these three performance indicators and sulfur values at each location.

As TU Electric enters its second decade of FGD operating experience, the progress of the first ten years summarized in this paper provides a framework to evaluate decisions on new and existing units. An achievement of 99% SO_2 compliance is a realistic target. Similarly, a Megawatt limitation factor below a target range of .5 to .75% is achievable but must be evaluated in terms of capital costs for redundancy of towers. The area with an immediate payback is O&M costs. Utilizing experience gained in achieving better limitation and compliance factors, TU Electric obtained more reliable process control enabling operating costs to become essentially fixed. Maintenance costs have been reduced to the bare bottom by upgrading when necessary or vital. Better equipment as developed will cost money. The payback of such an expenditure should be justified with an evaluation of "bottom line" performance parameters.

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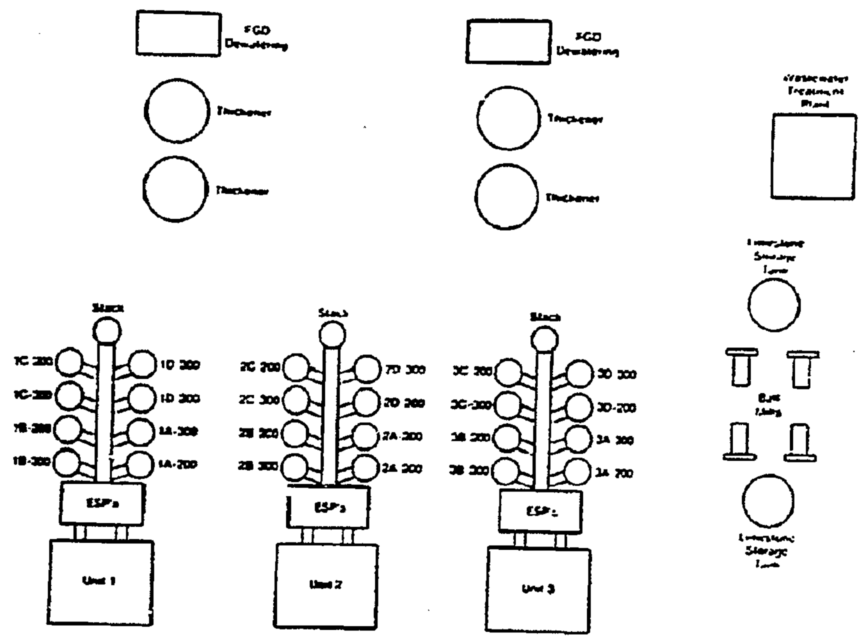
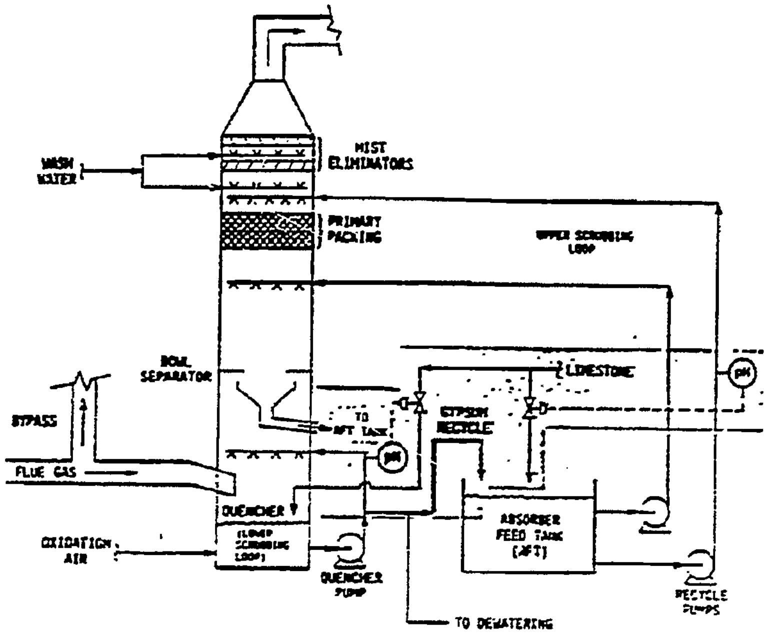
Fig. 1

MARTIN LAKE UNITS NO. 1, 2, & 3 SCRUBBERS

PLAN VIEW



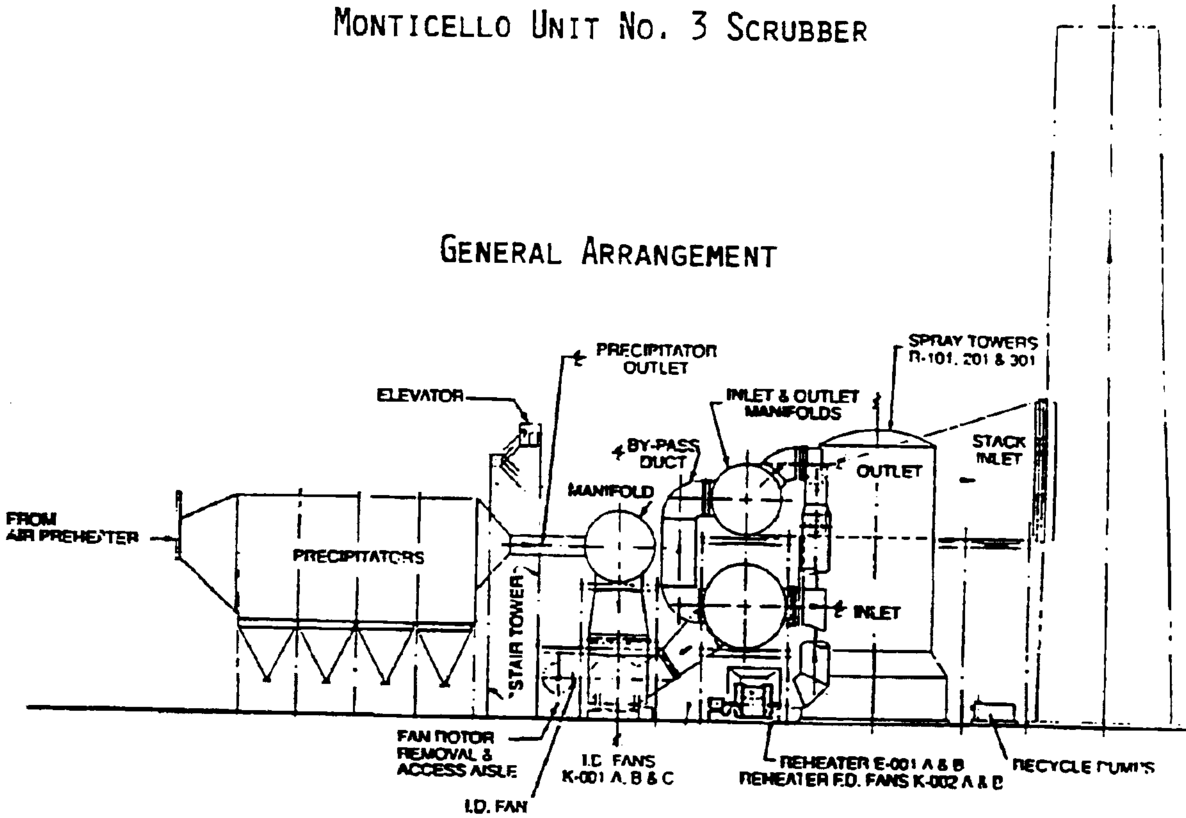


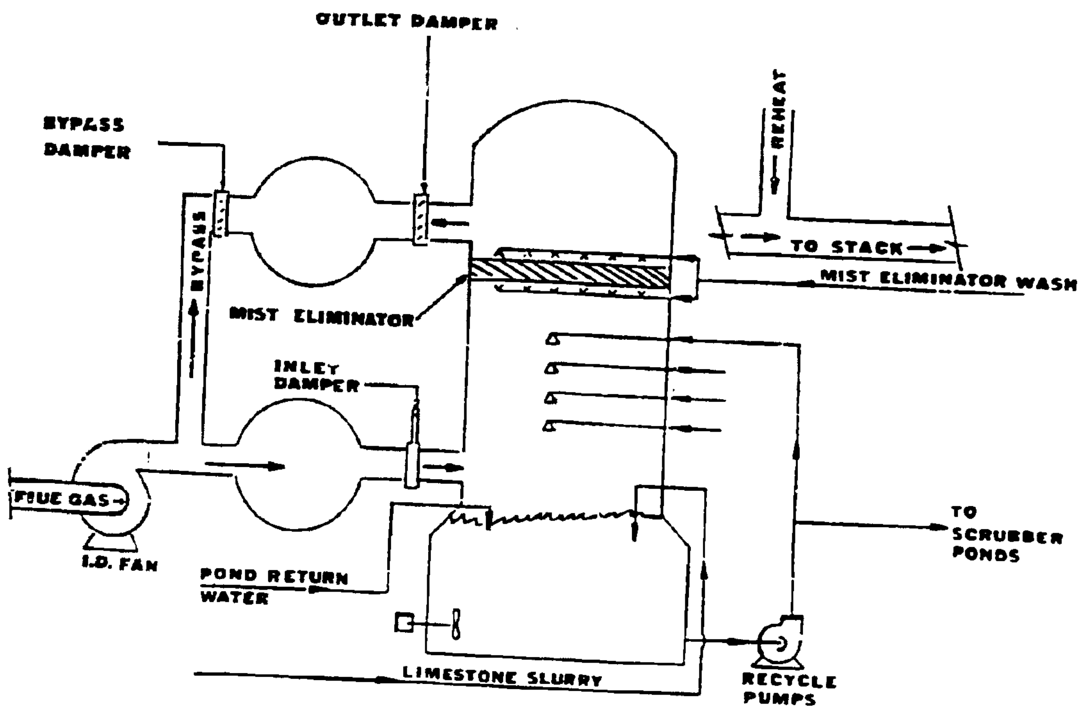
FGD SYSTEM SCHEMATIC

Fig. 2

MONTICELLO UNIT NO. 3 SCRUBBER

GENERAL ARRANGEMENT



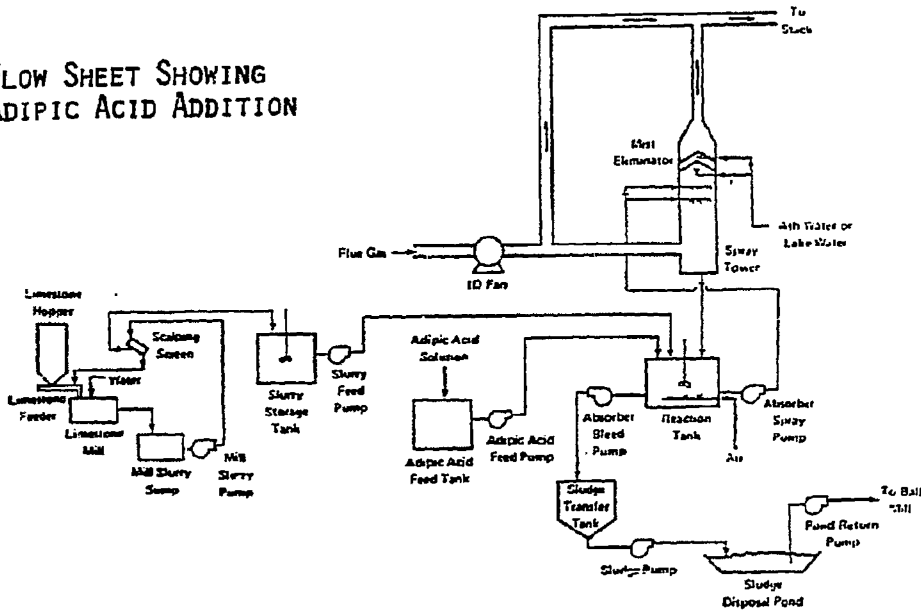


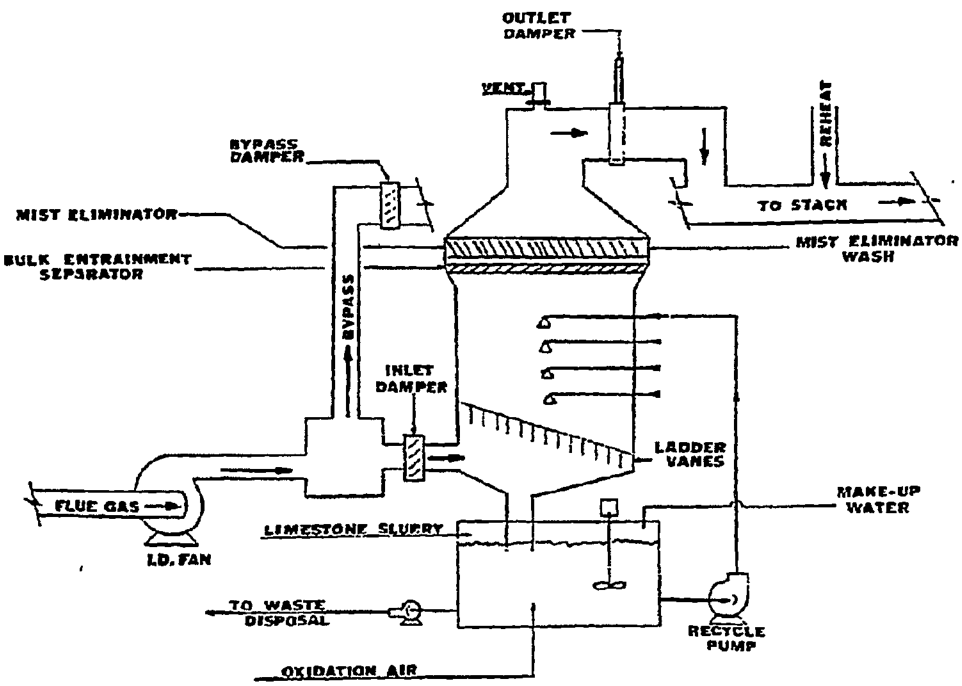
FGD SYSTEM SCHEMATIC

Fig. 3

SANDOW UNIT NO. 4 SCRUBBER

FLOW SHEET SHOWING ADIPIC ACID ADDITION

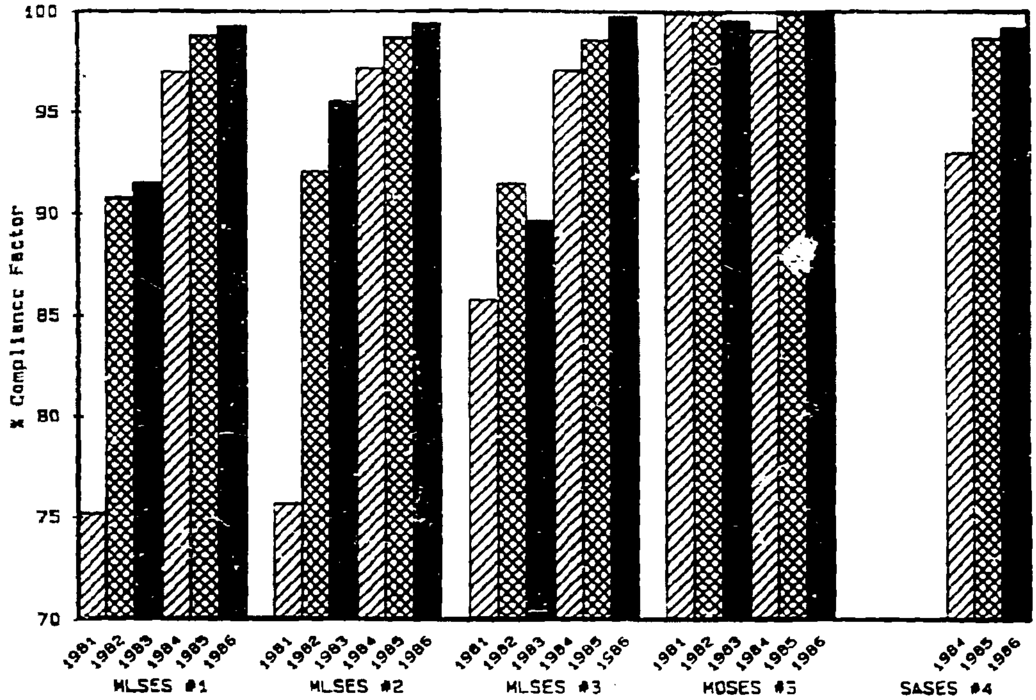




FGD SYSTEM SCHEMATIC

Fig. 4

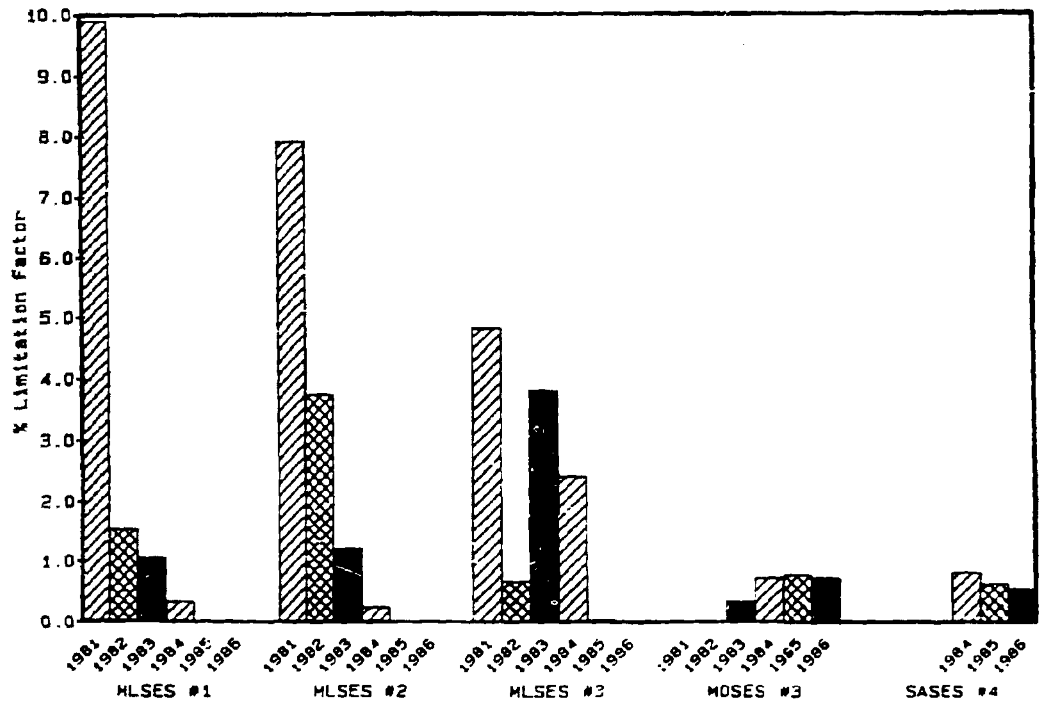
SCRUBBER COMPLIANCE FACTOR



$$\text{COMPLIANCE FACTOR} = \frac{\text{HOURS IN COMPLIANCE}}{\text{HOURS GENERATOR ON LINE}} \times 100$$

Fig. 5

SCRUBBER LIMITATION FACTOR



$$\text{LIMITATION FACTOR} = \frac{\text{FOH+EFOH}}{\text{HOURS GENERATOR ON LINE + FOH + EFOH}} \times 100$$

FOH = FULL UNIT OUTAGE ATTRIBUTED TO SCRUBBER
EFOH = EQUIVALENT UNIT OUTAGE ATTRIBUTED TO SCRUBBER (INCLUDES LOAD CURTAILMENT TO MAINTAIN COMPLIANCE)

Fig. 6

SCRUBBER COST PER UNIT ENERGY

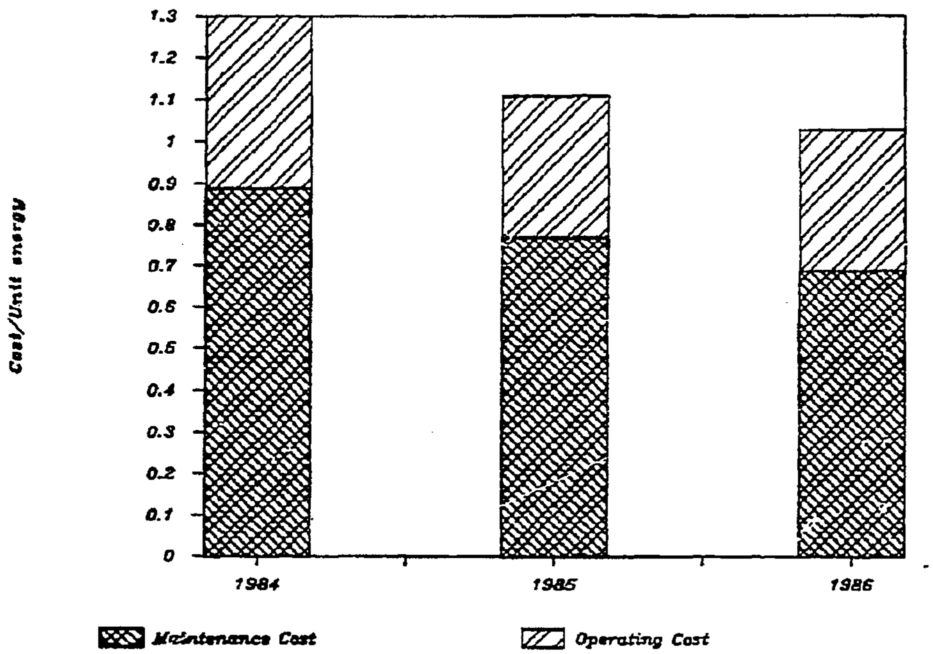


Fig. 7

		Martin Lake Units 1-3						Monticello Unit 3						Sandow Unit 4			
		1981	1982	1983	1984	1985	1986	1981	1982	1983	1984	1985	1986	1983	1984	1985	1986
PERCENT SULFUR	Max.	2.53	2.54	2.00	2.41	1.26	1.32	.68	.85	1.24	.90	.77	.94	1.25	1.90	1.65	2.06
	Min.	.42	.29	.44	.41	.51	.52	.36	.39	.48	.44	.28	.40	.98	.78	.78	.77
	Avg.	.97	.93	.93	.82	.75	.76	.52	.62	.68	.61	.58	.60	1.08	1.07	.98	.98
COMPLIANCE FACTOR		78.9	91.4	92.3	97.1	98.7	99.5	99.9	99.9	99.6	99.1	99.9	99.9	-	93.0	98.7	99.2
LIMITATION FACTOR		7.5	2.2	2.0	1.0	.04	.01	.04	.02	.35	.73	.78	.73	-	.82	.64	.56

Total O&M Cost - Avg. 5 Scrubbers			
	1984	1985	1986
Cost / Unit Energy	1.30	1.11	1.03