DOE/PC/92642--T1

CONSTRUCTION AND STARTUP EXPERIENCE FOR THE MILLIKEN FGD RETROFIT PROJECT

DE-FC22-93PC 92642 Dec. 1996

JAMES HARVILLA & MARK MAHLMEISTER
New York State Electric & Gas Corp.
Binghamton, New York

THOMAS BUCHANAN & CHRIS JACKSON
Parsons Power Group Inc.
Reading, Pennsylvania

JAMES WATTS
U.S. Department of Energy, P.E.T.C.
Pittsburgh, Pennsylvania

Abstract

Under Round 4 of the U.S. Department of Energy's Clean Coal Technology program, NYSEG, in partnership with Saarberg-Holter-Umwelttechnik, Consolidation Coal Company and Stebbins Engineering and Manufacturing Company, has retrofitted a formic acid enhanced forced oxidation wet limestone scrubber on Units 1 & 2 at the Milliken Steam Electric Station. Units 1 & 2 are 1950's vintage Combustion Engineering tangentially fired pulverized coal units which are rated at nominal 150 MW each and operate in balanced draft mode. The FGD system for Unit 2 was placed into operation in January 1995 and the Unit 1 system in June, 1995.

The project incorporates several unique aspects including low pH operation, a ceramic tile-lined cocurrent/countercurrent, split module absorber, a wet stack supported on the roof of the FGD building, and closed loop, zero liquid discharge operation producing commercial grade gypsum, and calcium chloride brine. The project objectives include 98% SO_2 removal efficiency while burning high sulfur coal, the production of marketable byproducts to minimize solid waste disposal, zero wastewater discharge and space-saving design.

The paper provides a brief overview of the project design, discusses construction and startup issues and presents early operating results. Process capital cost and economics of this design, procure and construct approach are reviewed relative to competing technologies.

INTRODUCTION

The Milliken Clean Coal Demonstration Project is one of nine projects selected for funding in Round 4 of the U.S. DOE's Clean Coal Demonstration Program. The project provides full-scale demonstration of a combination of innovative emission-reducing technologies and plant upgrades for the control of sulfur dioxide (SO_2) and nitrogen oxides (NOX) emissions from a coal-fired steam generator without a significant loss of station efficiency. The project incorporates several unique aspects including low pH operation, a ceramic tile-lined, cocurrent/ countercurrent, split module absorber, a wet stack supported on the roof of the FGD building, and closed loop, zero liquid discharge operation producing commercial grade gypsum and calcium chloride brine.

The project's sponsor is New York State Electric and Gas Corporation (NYSEG). Project team members include CONSOL Inc., Saarberg-Holter-Umwelttechnik (SHU), NALCO FuelTech, Stebbins Engineering and Manufacturing Co., DHR Technologies, Inc. and ABB Air Preheater. Project cofounders include NYSEG, CONSOL, Electric Power Research Institute, New York State Energy Research and Development Authority and Empire State Electric Energy Research Corporation. Parsons Power Group is the Architect/Engineer and Construction Manager for the flue gas desulfurization (FGD) retrofit portion of the project.

The overall project goals are:

- 98% SO_2 removal efficiency using limestone while burning high sulfur coal
- Up to 70% NOX reduction using the NOXOUT selective non-catalytic reduction (SNCR) technology in conjunction with combustion modifications
- Minimization of solid wastes by producing marketable by-products including commercial grade gypsum, calcium chloride, and fly ash
- Zero wastewater discharge
- Maintenance of station efficiency by using a high-efficiency heat-pipe air heater system and a low-power-consuming scrubber system.

The host site for the demonstration project is NYSEG's Milliken Station, located in Lansing, New York. Milliken Station has two 150-MWe pulverized coal-fired units built in the 1950's by Combustion Engineering. The SHU FGD process and the combustion modifications are installed on both units, but the NOXOUT process, Plant Economic Optimization Advisor (PEOA), and the high-efficiency air heater system are applied on only one unit.

The total cost of the project, including the three-year demonstration program, was budgeted at $158,607,807 with DOE contributing $45,000,000.

PROCESS OVERVIEW

SO_2 Removal

The SHU process (Figure 1) is the only developed wet-limestone FGD process designed specifically to employ the combined benefits of low-pH operation, formic acid enhancement, single-loop cocurrent/countercurrent absorption, and in situ forced oxidation. In the SHU process, the flue gas is scrubbed with a limestone solution in a cocurrent/countercurrent absorber vessel.

dg

DISCLAIMER

DISCLAIMER

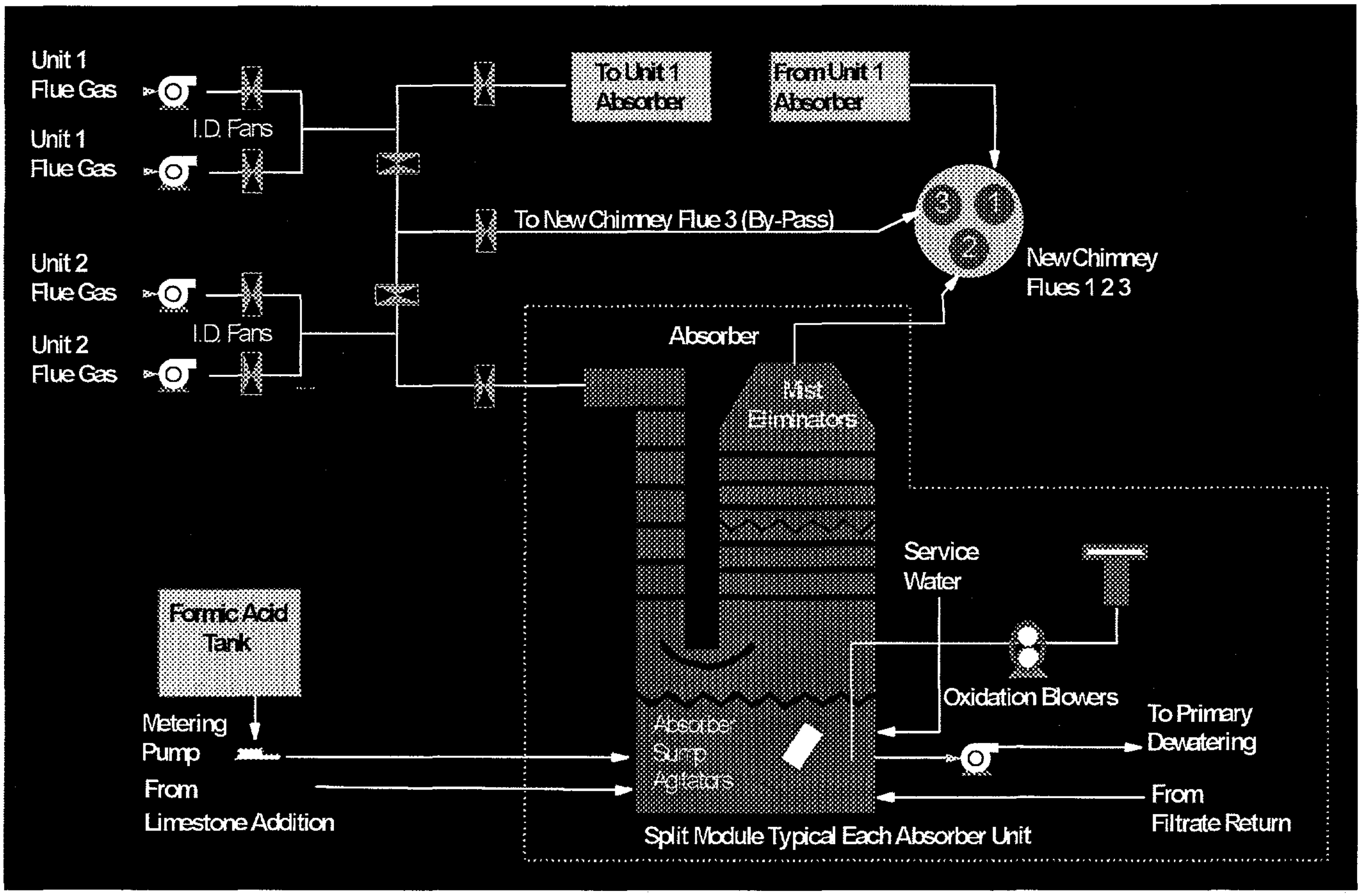


Figure 1. Simplified SHU Flow Diagram

The SHU solution is maintained at a low pH by adding formic acid, which acts as a buffer, to the absorber. Formic acid addition enhances the process in several ways, including better SO_2 removal efficiency with limestone, lower limestone reagent consumption, lower blowdown rate, freedom from scaling and plugging, higher availability, lower maintenance, production of wallboard grade by-product, and improved energy efficiency compared to conventional FGD technologies.

With operation at lower pH, the limestone reagent dissolves more quickly. This means that less limestone is needed, the limestone doesn't have to be ground as finely, and there is less limestone contamination of the gypsum by-product. Operation at lower pH results in more efficient oxidation of the bisulfite reaction product to sulfate. Less excess air is needed for the oxidation reaction and the gypsum crystals created are larger and more easily dewatered. Formic acid buffering improves SO_2 removal efficiency. Slurry recirculation rates are reduced, saving both capital cost and energy. Buffering provides excellent stability and ease of operation during load changes and transients. The process can tolerate higher chloride concentrations, reducing the amount of wastewater that must be processed. Finally, the potential for scaling of absorber internals is eliminated, resulting in reduced maintenance costs and improved availability.

The FGD process has been installed on both Units 1 and 2 with common auxiliary equipment. A single split absorber is used. This innovation features an absorber vessel divided into two sections to provide a separate absorber module for each unit. The design allows for more flexibility in power plant operations than does a single absorber while saving space on site and capital cost compared to two separate absorber vessels. The absorber shell construction is concrete, with an integral, cast in place ceramic tile liner. The tile has superior abrasion and corrosion resistance compared to rubber and alloy linings and is expected to last the life of the plant. In addition, the concrete/tile system is easily installed at existing sites where space for construction is at a premium, making it ideal for use in retrofit applications.

The absorbers use two-stage mist eliminators furnished by Munters. Whereas model DV 210 is used for the first stage in both absorber modules, the modules use two different second-stage designs. One absorber uses model DV-2130 and the other uses model T271. Model T271 is the vertical flow type tested by EPRI and commonly found in U.S. installations. DV-2130 is the Munters-Euroform V-shaped module design commonly used in European installations. The project will provide a side-by-side performance comparison of the two designs.

The design incorporates a new chimney erected on the roof of the FGD building, directly over the absorber vessel. Each absorber module discharges directly into a dedicated fiberglass (FRP) flue. The two FRP flues, provided by AN-COR Industrial Plastics, along with a common steel start-up bypass flue, are enclosed within a 40-ft diameter steel chimney. This design saves space on site and eliminates the need for absorber outlet isolation dampers, which are typically high maintenance items.

Limestone Preparation and Addition

Limestone is delivered to the station by truck. Space is provided on site for a 180-day inventory. The stone is reclaimed by front-end loader and transferred by belt conveyor to two 24-hr surge bins in the FGD building. The limestone is ground and slurried with clarified water (recycled process liquor) in conventional closed-circuit, horizontal, ball mill, wet-grinding systems (Figure 2) provided by Fuller. The 25% solids product is transferred by gravity to either of two 12-hour fresh slurry feed tanks. Redundant, continuous-loop piping systems are used to transfer the product slurry to the absorbers from the fresh slurry feed tanks. Two grinding systems are provided, each with a capacity of 24 tph. One mill, operating 12 hours per day, can support the process. Each system is provided with two sets of classifiers. This allows the production of slurry with two different particle size distributions, 90% passing through 170 mesh and 90% passing through 325 mesh. The coarser grind is used during normal operation with formic acid. The finer grind allows the system to be operated without formic acid. The limestone preparation/addition system can be aligned as two independent trains, effectively segregating Unit 1 and Unit 2 process streams. This feature enhances the flexibility of the installation for process evaluation purposes.

Gypsum Dewatering

A bleed stream of scrubber slurry is processed for recovery of high quality by-product gypsum and calcium chloride brine. Water is recovered and recycled back to the process. There is zero wastewater discharge from the process. The gypsum is dewatered to 8% surface moisture for delivery to customers in granular form. The absorber building has been designed for future addition of agglomeration equipment should market conditions require agglomerated product.

In the dewatering system (Figure 3) a bleed stream containing by-product gypsum solids is withdrawn from each absorber module by bleed pumps. The bleed streams are fed to primary hydrocyclones where the gypsum solids are concentrated to 25 wt%. The underflow from the primary hydrocyclones

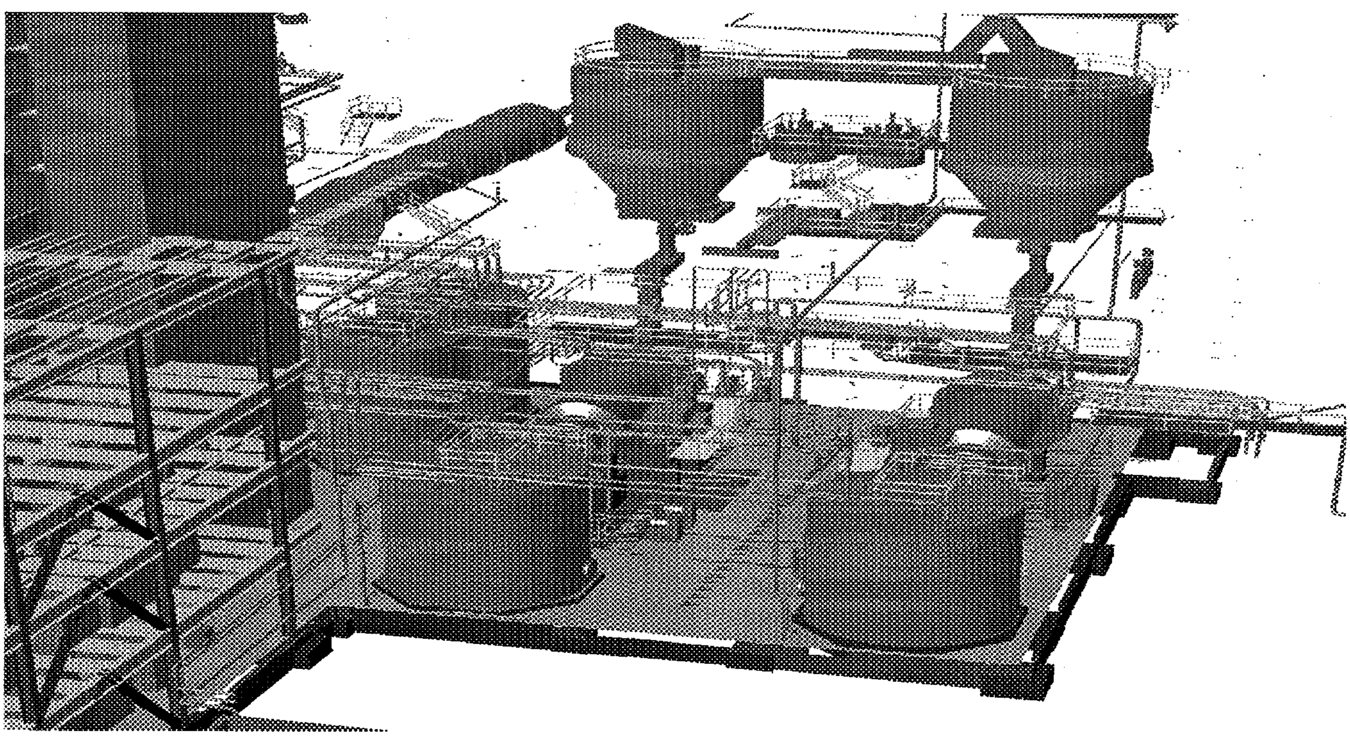


Figure 2. Limestone Preparation System

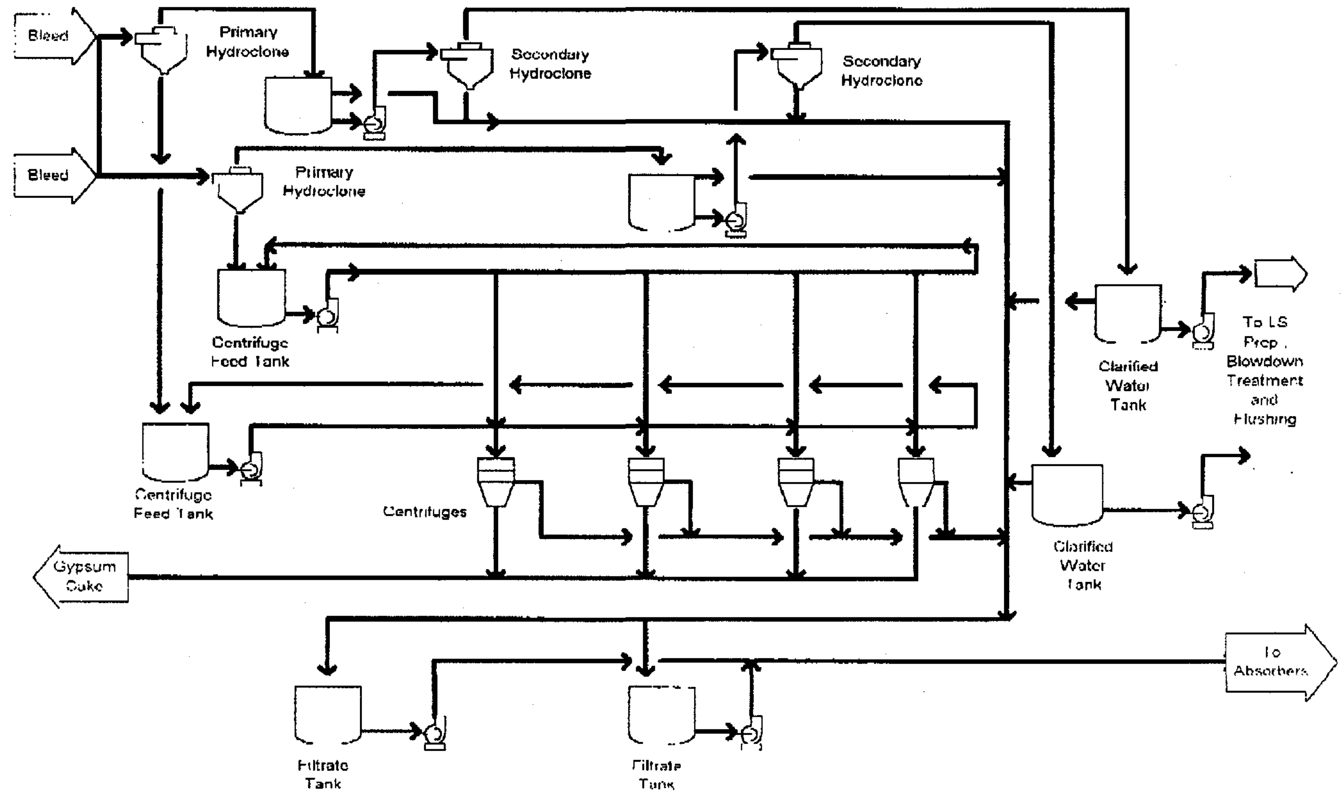


Figure 3.Gypsum Dewatering Flow Diagram

discharges to the centrifuge feed tanks. The overflow discharges to the secondary hydrocyclone feed tanks. Two primary hydrocyclone assemblies are provided by Warman. Each assembly can process the bleed from either or both absorber modules. The feed manifold of each hydrocyclone assembly has an internal partition which segregates the Unit 1 and Unit 2 bleed streams. This feature ensures that the feed rate to each individual hydrocyclone is constant whether or not the assembly is handling the bleed from one or both absorbers. In normal operation, the bleed from both absorbers is processed through one hydrocyclone assembly and the second assembly is a spare. If desired, both assemblies can be operated in parallel.

The gypsum solids from the primary hydrocyclone underflow are concentrated to 92 wt% by Krauss-Maffei vertical basket centrifuges. Four centrifuges are provided, three operating and one standby. The centrifuges are fed from either of two centrifuge feed tanks through continuously circulating feed loops. The rubber-lined centrifuges are batch operated and incorporate a washing step designed to achieve a residual chloride concentration of less than 100 ppm. The system is configured to allow segregation of the Unit 1 and Unit 2 liquid streams. The centrate is returned to the absorbers through the filtrate tanks. The gypsum solids are transferred by belt conveyor to an on-site storage building. Gypsum in the 5000-ton capacity storage building is reclaimed by front-end loader and trucked from the site.

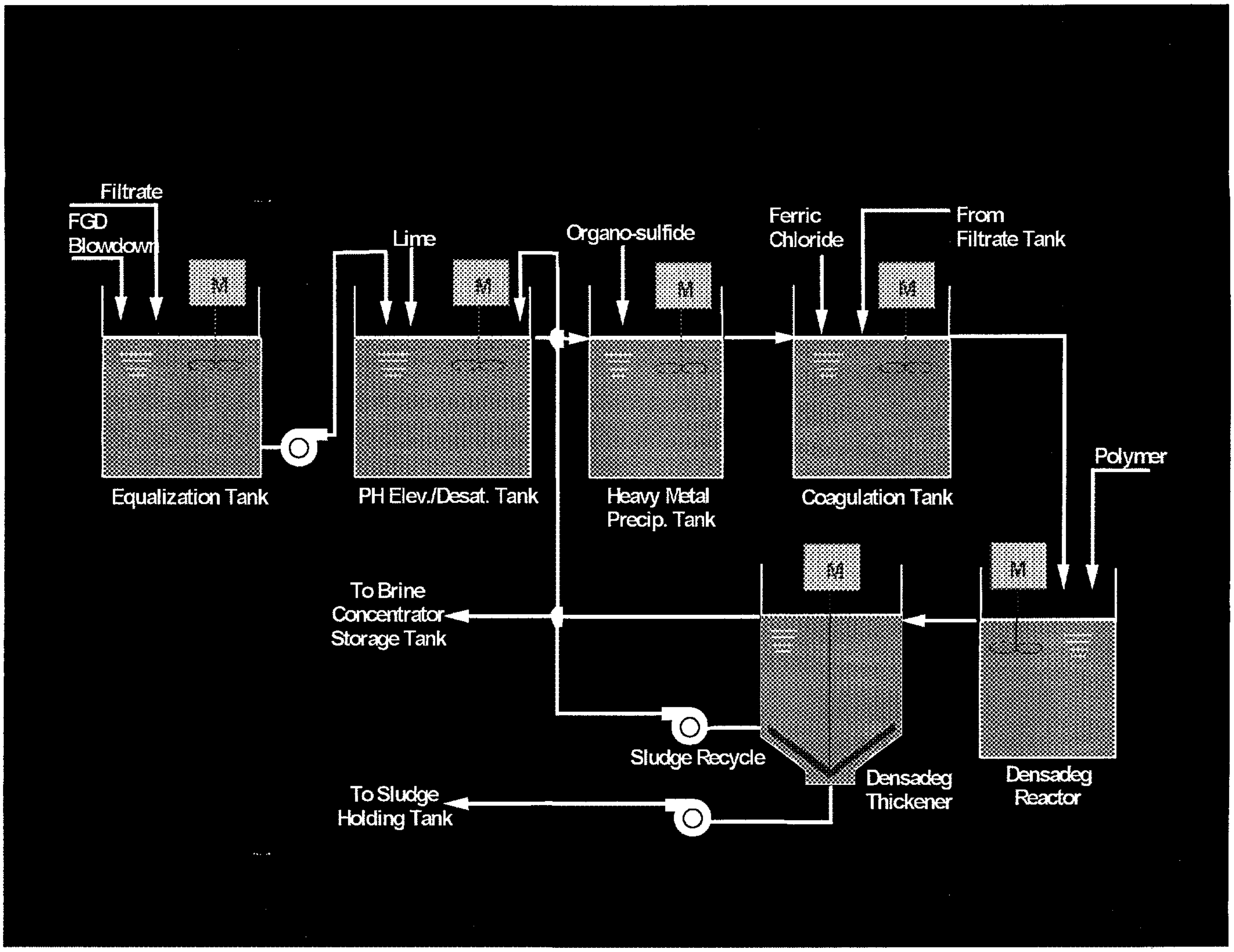
A portion of the overflow from the primary hydrocyclones is processed by the secondary hydrocyclones for use as clarified water for limestone preparation, system flushing, and blowdown to the FGD wastewater treatment system. Gypsum solids in the underflow from the secondary hydrocyclones and the balance of the primary hydrocyclone overflow are returned to the absorbers via the filtrate tanks. Two secondary hydrocyclone assemblies are provided by Warman, one dedicated to each primary hydrocyclone assembly, maintaining the capability of segregating the Unit 1 and Unit 2 process streams.

FGD Blowdown Treatment

The FGD Blowdown Treatment System consists of two subsystems, the pretreatment system, furnished by Infilco Degremont Inc.(IDI), and the brine concentration system, furnished by Resources Conservation Co.(RCC). The project is the first demonstration of the production and marketing of FGD by-product calcium chloride.

The pretreatment system (Figure 4) removes suspended and dissolved solids from the blowdown stream prior to the brine concentration process. The pretreatment process consists of the following steps:

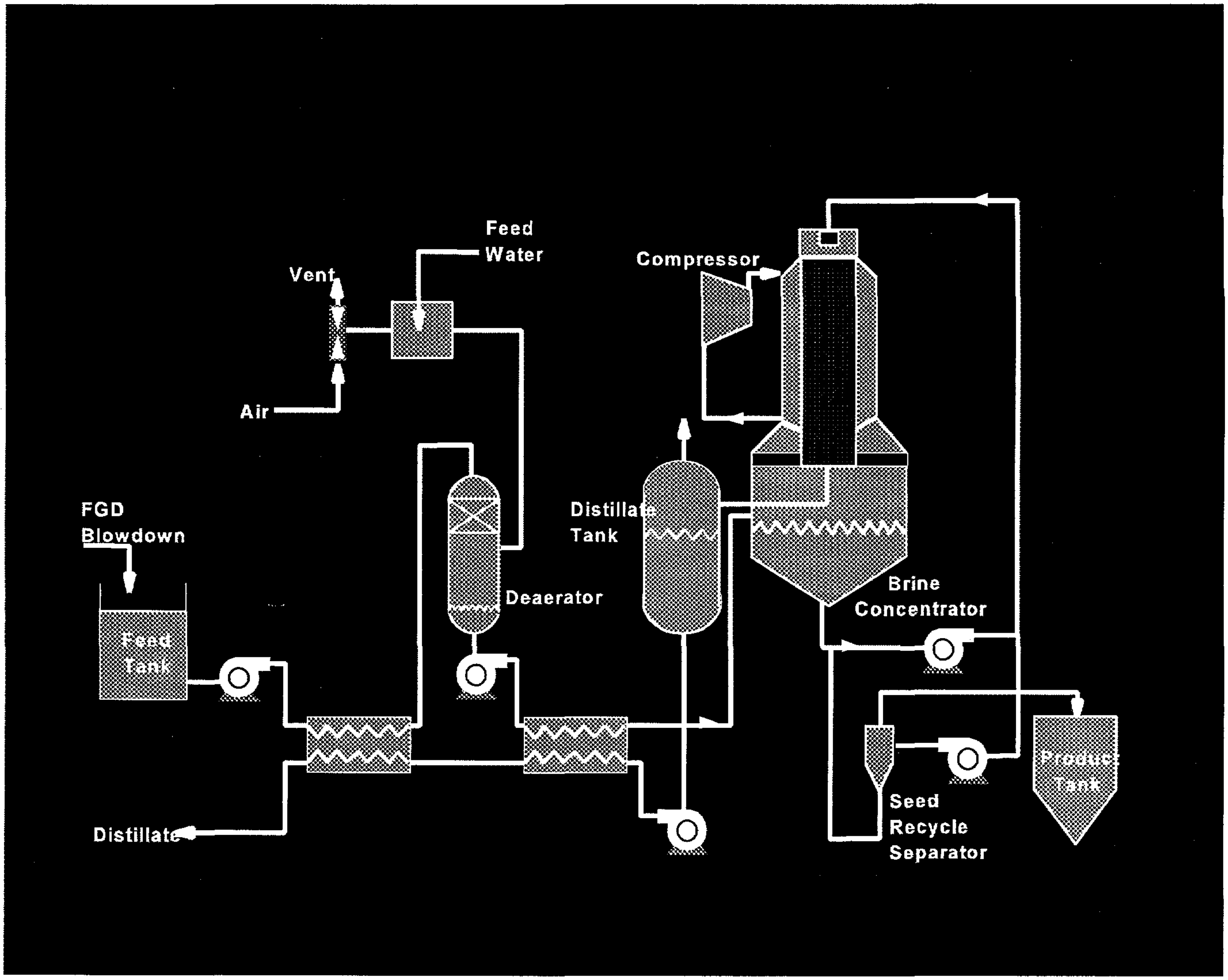
- An agitated equalization tank to balance the FGD wastewater composition and flow.



- pH elevation, calcium sulfate desaturation and magnesium hydroxide precipitation using lime. By elevating the pH, most heavy metals are removed. In particular, the high pH leads to precipitation of magnesium hydroxide, leading to a purer calcium chloride salt product. Sludge is recirculated from the downstream clarifier to aid the desaturation process.
- Secondary precipitation of heavy metals as more insoluble organosulfides using the organosulfide TMT.
- Coagulation with ferric chloride.
- Dosing of flocculant to the reactor of the DensaDeg unit to improve sedimentation.
- Flocculation/sludge densification, thickening, and final clarification in the DensaDeg unit. The DensaDeg is a three-stage unit comprising a solids-contact reaction zone, a presettler-thickener, and lamellar settling tubes in the upper part of the thickener. The water entering the clarification zone has a very low solids content and the lamellar tubes serve only to catch fugitive particles carried over. Water leaving this zone has less than 20 ppm solids.
- Excess sludge withdrawal, conditioning with lime, and dewatering with a plate and frame filter press. The addition of lime in the sludge holding tank aids the dewaterability of the sludge, allowing a drier cake to be formed, and also helps stabilize the metal hydroxides.

The brine concentration system (Figure 5) processes the effluent from the pretreatment system through a vapor-compression type falling-film evaporator, producing a very pure distillate that is recycled to the FGD system as process makeup water. The system's by-product is calcium chloride brine suitable for use in dust control, soil stabilization, ice control, and other highway construction related purposes.

The pretreated FGD blowdown is conditioned with hydrochloric acid and an inhibitor for scale prevention. It is then preheated, deaerated, heated to near boiling, and fed to the evaporator sump where it mixes with recirculating, concentrated brine slurry. The slurry is pumped to the brine concentrator (BC) condenser floodbox where it is distributed as a thin film on the inside walls of titanium tubes. As the slurry film flows down the tubes, the water is evaporated. The resulting steam is drawn through mist



eliminator pads to the vapor compressor, which raises its saturation temperature to above the boiling temperature of the recirculating brine. The compressed steam is then introduced to the condenser where it gives up its heat of vaporization (to heat the thin film in the inside of the tubes) and condenses on the outside of the tube walls. This condensate is collected in the distillate tank, cooled by heat exchange with the feed stream, and returned to the FGD system. As the falling film evaporates, calcium sulfate begins to crystallize. The calcium sulfate seed crystals provide nucleation sites to prevent scaling of the tubes. Control of the concentration of both suspended and dissolved solids in the evaporator sump is critical to prevent the precipitation of secondary salts and the resultant scaling of the evaporator tubes. A side stream of recirculating brine is processed by a hydrocyclone. The underflow is returned to the BC sump. The overflow is either recirculated to the brine concentrator or diverted to the product tank, based upon its dissolved solids concentration. A second side stream of recirculating brine is diverted to the product tank to control the concentration of suspended solids. The 33% brine product is then cooled and transported to market by truck.

CONSTRUCTION

Engineering and design work for the project began in January 1992. Construction started in April 1993, and was completed in time to begin scrubbing the first unit in January, 1995.

Schedule

As with most FGD retrofit projects, running a major construction project on a site shared with an operating unit posed several construction coordination challenges. One of the major drivers behind the construction plan, in addition to DOE's commitment to be ready to begin the demonstration program in June 1995, was the desire to use existing unit scheduled outages for tying in the FGD systems. This strategy avoided the project's causing the station to lose generating time and the associated revenue. Unit 2 was scheduled for a maintenance outage in late 1994 and Unit 1 in spring 1995. Since only a partial bypass is provided around the scrubbers, once a unit was tied in, the FGD system had to be operational.

Absorber Module Construction

Meeting the Unit 2 outage schedule meant installing mechanical equipment as well as piping, the absorber vessel, and the roof-mounted chimney during the upstate Finger Lake region winter. It was therefore essential that the FGD building be erected and enclosed by January 1994. Stebbins' unique construction method, which uses the Stebbins tile liner as the formwork for the concrete pours, limits the height of each pour to about one foot. Accordingly, 33 weeks were scheduled for erection of the 108-ft tall absorber vessel. This meant that the building steel had to be erected in parallel with the absorber. To accommodate the associated safety issues, the initial vessel erection was done on the second shift. The building was enclosed in time to allow mechanical work to proceed without major disruption from the unusually severe winter weather.

Chimney Erection

International Chimney mobilized on site in December 1993, and began erecting the stack in January 1994. The 140-ft tall, 40-ft diameter steel shell was fabricated on site in 10-ft sections, lifted into position by the 350 ton DeMag, using a 420-ft boom, and welded in place. The 12-ft diameter, 227-ft tall FRP flues were shop fabricated in 40-ft spools, lifted into the shell with the crane and attached with bell/spigot FRP butt welds. The stack was topped out in May, in time to make way for erection of the limestone and gypsum conveyors.

Limestone and Gypsum Conveyors

The conveyors, furnished by FMC, were prefabricated in tubular galleries to minimize field erection time. The 308 ft long 8'8" diameter limestone conveyor gallery came in eight spool sections, ranging in length from 15 to 50 feet. The 11 ft diameter gypsum conveyor came in a single 55 ft spool. The sections were lifted in place with the DeMag and welded together. Erection of the conveyors was scheduled to take five months.

STARTUP ISSUES

Startup Program Overview and Schedule

The startup schedule was developed to provide a well organized and logical sequence of events necessary to meet the scheduled Unit 2 FGD system on-line date of January 17, 1995. Due to the two train design utilized for redundancy, most of the support equipment for the Unit 1 module was also ready at this time. The actual date for gas to Unit 1 was not until June. All support equipment systems were started as early as possible. These included: steam for plant heating, station electric supplies, service and instrument air, and service water. Major systems were scheduled depending upon the completion of associated construction activities and the levelization of startup manpower.

The major construction activity influencing startup sequencing was the completion of the absorber vessel. Equipment not included in this category was limited to the ball mills and material handling. The waste water treatment and brine concentrator systems were started last to allow the dissolved solids in the scrubber to build up after gas treatment began. Table 1 provides the startup sequence of the scrubber:

Table 1

Scrubber Start-Up Sequence

1	Sumps and Drains	7/28/94
2	Process Water	7/29/94
3	Make-Up, Wallwash and Quench	8/5/94
4	Limestone Handling	8/12/94
5	Filtrate	9/26/94
6	Fresh Slurry	10/4/94
7	Mill Product	10/10/94
8	Weigh Feeder	10/10/94
9	Slurry Bleed	10/11/94
10	Clarified Water	10/12/94
11	Gypsum Handling	10/26/94
12	Ball Mill Lube System	11/2/94
13	Absorber Agitators	11/9/94
14	Centrifuges	11/15/94
15	Absorber Recycle Pumps	11/17/94
16	Formic Acid	11/17/94
17	Mist Eliminators	11/29/94
18	Flue Gas (CEMs, Seal, Air, Dampers)	12/9/94
19	Emergency Quench	12/13/94
20	Oxidation Blowers	1/7/95
21	Ball Mills	1/10/95
22	Blowdown Pretreatment	3/27/95
23	Brine Concentrator	7/15/95

SO_2 Removal

Quench System Modifications - During the initial operation of the Unit 2 scrubber module, control problems were experienced with the normal and emergency quench systems. Process water flow to the normal quench system was controlled by monitoring three thermocouples installed above the first recycle spray header. The output from the thermocouples was not consistent, causing the quench flow set-point to fluctuate and the generation of nuisance alarms. Since under most conditions the absorber make-up requirements are higher than minimum process water flow rates in the system, the normal quench control was changed to be a constant flow. Utilizing a constant flow also tends to minimize the risk of quench spray nozzle pluggage.

The emergency quench system is designed to maintain a maximum gas temperature at the third recycle spray header elevation on the cocurrent side. This protects the rubber lined structural support steel and the fiberglass recycle headers located downstream. The headers and support steel located above this elevation are constructed of alloy C-276. Similar to the normal quench control system, three thermocouples located above the third recycle spray header were designed to monitor gas temperature and control the emergency quench sprays. During a high temperature condition, the emergency diesel fire pump will start and the emergency quench valve will open. Operation of this system for prolonged periods will cause water balance problems.

Shortly after start-up we began to observe erratic temperature indications from the three thermocouples. Due to the critical nature of the system, we temporarily installed three additional thermocouples at the same elevation. By comparing results from all six instruments, we concluded the problem was being caused by localized impingement of recycle spray. The problem was solved by installing longer elements to move away as far as practical from the nearest spray nozzle and still protect the module. Inspections of rubber surfaces at the third spray elevation in the module following this change have shown no thermal effects.

pH Measurement Modifications - To protect the module internals and maintain process water balance, one of the first two recycle headers on the cocurrent side must be in service at all times. The module pH sample is obtained by taking a slip-stream from one of these headers and routing it through a sample box. The box was constructed of flake-glass lined carbon steel. Due to the highly abrasive and corrosive nature of the slurry, leaks developed in the sample box after the first six months. We have since replaced the box with an inverted U-tube and orifice. The fix has been operating satisfactorily for two months. Since the U-tube was constructed of PVC piping, we plan to replace it in 1996 with abrasion resistant FRP pipe.

Solids Accumulation in Absorber Sump - Following the first 40 days of operation a short scheduled outage was planned for the scrubber coinciding with the removal of the fine screens on the turbine. When the module was inspected, an accumulation of solids was discovered along the north wall of the module. Approximately 25 tons of material was removed during the outage. The accumulation appeared to be layered indicating the cause was not a one-time event such as a shutdown or startup. We plan to address this problem by speeding up the impellers on the absorber agitators. New sheaves will be installed during 1996.

Solids in Clarified Water - Due to the use of rubber lined components we developed a problem with small pieces of rubber circulating in the system. The rubber and other debris would eventually show up in the clarified water system. The scrubber system also has some areas that are open which may allow foreign materials to enter the process, such as hydroclone underflow launders. The original pump strainers were cumbersome to remove and clean and were only intended for use as startup strainers. To make it easier to remove materials from the system, we installed permanent PVC basket strainers on the clarified water pump suctions.

Limestone Preparation

The startup of the limestone grinding system required minimal field changes. The system is designed to accommodate two grind sizes for the two operating scenarios, with formic (90% passing 170 mesh) and without formic acid (90% passing 325 mesh). The mill product pumps used to feed the hydroclones to classify slurry solids are variable speed drive and are controlled by maintaining a set pressure on the inlet to the hydroclones. The operating pressure was adjusted to provide the most efficient grinding recirculation rate for the two grind sizes. Following the initial grinding and solids particle size analysis, the number and size segregation of balls was adjusted. This system has been operating satisfactorily since the initial start-up. No sizing adjustments were made to the apex or vortex finders in the hydroclones.

Gypsum Dewatering

Optimization of Hydrocyclone Vortex Finder and Apex - The primary and secondary hydroclones were supplied with rubber vortex finders and apexes. Proper sizing of hydroclone internals is critical to the operation of the scrubber system. The underflow from the primary hydroclones is used as the centrifuge feed slurry. The feed slurry flow rate and percent solids are important. The centrifuge feed loop relies on the capacity of a relatively small feed tank. If the feed system does not match the centrifuge required feed rate, the centrifuge will trip during a feed step. The overflow from the secondary hydroclones is used as clarified water for grinding and equipment flushing. If the clarified water flow rate and percent solids are not optimized, the grinding rate and flushing efficiencies will be affected. Since particle size distribution of gypsum crystals is somewhat site-specific, the final apex and vortex finder sizes were to be determined in the field.

Utilizing rubber internals enabled us to adjust performance in the field by clamping down on the apex. Initial field results led us to believe that we had a serious problem with hydroclone performance. However, a detailed site evaluation was completed including time averaged sampling of the overflow and underflow from the primary hydroclones. The problem was subsequently traced to the lab work. The lab was not accounting for the increased liquor density caused by dissolved solids, which resulted in high suspended solids calibrations of density meters. Based on the results of the site evaluation the vortex finder and apex sizes were finalized, enabling to system to perform as intended.

Centrifuge Cycle Optimization - Operation of a centrifuge cycle requires the following steps: ramp up, feed, cake wash (to remove chlorides), spin (dewatering at constant speed), ramp down, and cake peel. A heel rinse and basket rinse may also be required. Since the centrifuge operates in a batch process, the key to successful operation is to achieve the required residual moisture and chlorides in the gypsum cake with the fastest cycle times. Each step of the process is controlled by different parameters. Since the dewatering properties are dependent upon the site-specific crystal structure, the cycle parameters were optimized in the field.

The time to ramp up and down is set by output of the variable frequency drive unit. The feed time was optimized by controlling the feed density to a minimum and setting flow rates with manual concentric reducing valves. The cake wash was set by monitoring the residual chlorides in the cake as a function of wash time. Similarly, the spin time was set by monitoring the residual cake moisture as a function of spin time. The peel time was set by adjusting the knife travel speed and monitoring the discharge conveyor loading.

Modifications of Centrifuge Feed Piping - During initial operation of the centrifuges, the residual chloride level in the cake was not consistently below the 100 ppm level necessary for use by wallboard manufacturers. The cause was traced to a drain problem with the feed piping. A small amount of feed slurry was draining into the centrifuge during the cake wash step. The fix to this problem was to install drain valves in the horizontal section of feed pipe closest to the centrifuges. When the feed step is complete, the valves open. The residual chlorides in the gypsum cake have been consistently below 100 ppm since the valves were installed.

FGD Blowdown Treatment

Blowdown Pretreatment - The wastewater treatment system was started in April, 1995. The performance testing of the system was being conducted as this paper was being prepared in January 1996. We have had good operating experience with the system. The only changes made to the system following startup were relatively minor. The filtrate drain from the filter press was re-routed away from the equalization tank to maintain a more consistent raw feed. A magnetic flow meter and trap on the overflow from the thickener were moved to eliminate a gravity flow problem.

Brine Concentrator - The brine concentrator was started in July 1995. We are currently working to optimize system performance and have not yet conducted an acceptance test.

OPERATING RESULTS

Demonstration Testing Program

The demonstration testing program planned for the FGD System, summarized in Table 2, is designed to characterize the performance of the SHU FGD process. The testing program will be conducted over a period of 36 months. The goals of the program are to demonstrate the effectiveness of the process at several operating conditions and to demonstrate the system's long term reliability and performance. Typical evaluations will include SO_2 reduction efficiency, power consumption, process economics, load following capability, reagent utilization, by-product quality and additive effects.

Unit 1 will be operated continuously at the design conditions while parametric tests are performed on Unit 2 to define the performance limits of the FGD system. Because they are nearly identical modules, Unit 1 will provide a baseline while the parametric tests are being performed as well as serving as a long-term test. The parametric tests are set up to study the effects of formic acid concentration, L/G ratio, mass transfer, coal sulfur content and flue gas velocity on scrubber performance. Although load following capability will be monitored, load will not be a controlled variable. As much as possible, load changes during the parametric testing period will be handled by Unit 1 in order to keep Unit 2 at full load. The same coal will be fed to both units simultaneously. The chloride content will not be a controlled variable. At the design bleed rate chloride level is expected to stabilize at about 40,000 ppm Cl- by weight when burning a 0.1 wt % chlorine coal. Limestone utilization will be held constant at the design level except for a few FGDPRISM Model calibration runs. A list of process variables to be measured is shown as Table 3.

Table 2

Miliken Project Test Plans FGD Process Testing

Process	Variable	Variable Range	Goals
SHU	Sulfur Content	1.6% to 4.0%	- 95% to 98% SO_2 removal - 95% reliability - Minimum energy consumption for base coal - Determine impact of variables on $_2$ removal, gypsum quality and chloride brine quality when operating with design coal - Determine impact of FGD on net plant heat rate - Confirm calcium use and formic acid makeup rate
	Formic Acid Concentration	0, 400, 800, 1600 ppm	
	Combination of Spray Header	Various Spray Header Combinations	
	Gas Velocity in Cocurrent Section of Absorber	18 to 22 fps	
	Limestone Grind Size	90% - 170 mesh and 90% - 325 mesh	

Table 3
Process Variables to be Measured

Sample Location	Stream Type	Flow	Temp	SO_2 Content	P	Formic Acid Conc.	Pressure	pH	Density	Belt Speed	Cl Concentration	Level Indicator	On/Off Indicator	Moisture
Flue Gas from ID Fan to Absorber (500)	Gas	C	C	C										
Flue Gas in Chimney (501)	Gas	C		C										
Pressure Across Absorber (500) & (511)	Gas				C									
Compressed Oxidation Air to Absorber (505)	Gas		C				C							
Formic Acid to Absorber (504)	Liquid	C				P								
Recycle Scrubber Slurry (503) & (502)	Slurry					D	C	C						
Limestone Slurry from Slurry Tank (506)	Slurry						C							
Total Process Water to System (514)	Liquid	C					C							
Gypsum Slurry to Dewatering (600)	Slurry								C					
Limestone Feed Belt (400)	Solid									C				
Clarified Water to Mills (402)	Liquid	C					C							
Gypsum from Vacuum Filter to Storage	Solid									C			C	C
Clarified Water to Blowdown Treatment (700)	Liquid	C									D			
Filter Cake Wash Water (601)	Liquid	C	C				C							
All Process Tanks	Liquid											C		
Mist Eliminator	Gas				C									

C = Continuous Monitoring D = Daily Analyses P = Periodic

Test Parameters

Coal Sulfur Content - The plant design is based on a nominal coal sulfur content of 3.2 wt %. The project will use Pittsburgh seam coal. The coal sulfur content will be varied over a range of 1.6 to 4.0 wt % using at least three different coals. Tests will be performed using the lower sulfur coal first, followed by the design coal, and conclude with a short period using high-sulfur coal. The high-sulfur coal testing will be done on Unit 2 during a scheduled outage on Unit 1 because the equipment for dewatering and reagent preparation is not designed to handle the output of both units simultaneously using high-sulfur coal. Parametric tests will not be performed using high-sulfur coal but the process will be operated at optimum conditions based on the results of parametric tests using the design coal and FGDPRISM modeling results. The purpose of using high-sulfur coal is to demonstrate the operability of the process using 4% sulfur coal, not to determine the effect of operating parameters on performance.

Formic Acid Concentration - The process design is based on a certain concentration of formic acid in the scrubber slurry. Testing will be conducted at concentrations ranging from 0 to 150% of design. Ideally, in this type of testing program, all parameters should be randomized; however, the large capacity (270,000 gal) in the scrubber sump makes it impractical to frequently increase and decrease the formic acid concentration. Therefore, the program is set up in blocks of tests in which the formic acid concentration is kept constant for long periods of time (4 to 25 days). Each block of tests will be conducted in order of increasing formic acid concentration, because it takes substantially more time to lower the concentration than to raise it.

Limestone Grind Size - The design limestone grind size is 90%<170 mesh when using formic acid and 90%<325 mesh using no formic acid. The design grind size limestone will be used for all but a few test runs which will be done to observe the effects of grind size on performance.

Spray Header Combination-L/G Ratio - There are four cocurrent spray headers and three countercurrent spray headers in each SHU module. The spray headers operate in an on/off mode, i.e., there is no flow control on the headers. The scrubber L/G ratio is varied by changing the number of spray headers in operation. The process design calls for operation of five spray headers to achieve 95% SO_2 removal and all seven headers to achieve >98% SO_2 removal. At least two of the seven headers should be operating at all times. In addition, at least one of the top two headers on the cocurrent side must be operating at all times in order to protect vessel internals from over temperature. Parametric testing will include operating various combinations of spray headers in the cocurrent and countercurrent sections to determine the combination that provides the best SO_2 removal performance and lowest scrubber energy consumption. For each combination, the uppermost headers will be used. For each test coal, the pressure drop and SO_2 removal will be measured for each spray header combination used. The gypsum crystal morphology and formic acid consumption rate will be determined for selected spray header combinations using the design coal only.

The results of these tests will also be used to determine the mass transfer coefficients individually for the cocurrent and countercurrent sections. The results from tests with all countercurrent sprays turned off will be used to determine the mass transfer in the cocurrent section. The mass transfer in the countercurrent section will be determined by comparing these results with results from tests in which countercurrent sprays are operating.

Gas Velocity in the Cocurrent Scrubber Section - Tests at higher than design gas velocity will be performed on the Unit 2 scrubber by shunting some of the gas flow from Unit 1 to the Unit 2 scrubber. The purpose is to provide data on high gas velocity scrubbers. These tests will be performed using two formic acid concentrations (0 and design) and two coals (lower sulfur coal and the design coal). The pressure drop and SO_2 removal will be measured for several spray header combinations. The gypsum crystal morphology and formic acid consumption rate will be determined for selected spray header combinations while using the design coal.

Test Description

Tests Using Design Gas Velocity-Lower Sulfur Coal - This set of tests was conducted in November of last year. All of the possible spray header combinations were used for the tests using design gas velocity, design limestone grind size, and lower sulfur coal. Each test was repeated, giving 28 tests total at each formic acid concentration. These tests were run in random order at constant formic acid concentration. In addition, two tests were run at each formic acid concentration using an alternative grind size. The effect of grind size will be determined by comparing the results of these tests with the results of tests using the design grind size at the same header configuration and formic acid concentration. Each test was scheduled for eight hours. Pressure drop and SO_2 removals were measured after SO_2 levels had lined out. Gypsum crystal morphology was not be characterized in this series of tests.

Tests Using High Gas Velocity-Lower Sulfur Coal - This set of tests was conducted in November of last year. These tests were performed using no formic acid and the design formic acid concentration. A minimum of four total spray headers were in operation at all times. Five of the tests were repeated, giving thirteen tests total. The tests were run in random order using the design limestone grind size. SO_2 removal was measured. Alternative grind sizes were not be tested. Gypsum crystal morphology was not be characterized.

Tests Using Design Gas Velocity-Design Sulfur Coal - Fewer spray header combinations will be tested using the design coal because it is not a compliance coal. Since low L/G ratios might not remove enough SO_2 to keep the station in compliance, at least four spray headers will be operating at all times. If SO_2 removal drops to unacceptably low levels during a test, that test will be terminated and compliance performance will be reestablished before proceeding to the next test. Measurements and sampling during each test will include SO_2 removal, pressure drop, gypsum crystal morphology (particle size distribution, sulfate/sulfite ratio, and SEM micrographs), gypsum samples for wallboard evaluation, calcium and sulfur balances, formate consumption rate, and O_2 consumption for oxidation. Sampling will begin after 10 turnovers (4 days) have passed to insure solid phase lineout. The coarser (<170 mesh) grind-size limestone will not be tested without formic acid because of the danger of not reaching sufficient SO_2 removal for compliance. The limestone grind size is not something that can be changed quickly if higher SO_2 removal is needed. The finer (<325 mesh) grind size will be used for one test at each formic acid concentration. Data for FGD system performance guarantee verification will also be collected during this period.

Tests Using High Gas Velocity-Design Sulfur Coal - The same tests that were run using the low-sulfur coal at high gas velocity will be run using the design coal. If SO_2 removal drops to an unacceptable level during a test, that test will be terminated and compliance performance will be reestablished before proceeding to the next test. Alternative grind sizes will not be tested. SO_2 removal will be measured. Gypsum crystal morphology will not be characterized.

In addition to the above program, a short series of tests will be conducted in which the process pH will be varied from 4.5 to 5.5; these tests will be performed to calibrate the FGDPRISM model. The process operating conditions for these calibration tests will be set by EPRI after the model is programmed. These tests will be performed with no formic acid after completion of the test block using design gas velocity, lower sulfur coal and no formic acid.

Preliminary Test Results - Low-Sulfur Coal

SO_2 Removal - Parametric testing utilizing a 1.6% sulfur coal was completed on Unit 2 in November 1995. Test results were being reviewed during the preparation of this paper in January 1996. Due to the nature of a cocurrent/countercurrent scrubber, test results will vary depending upon the combination of spray headers in service. Therefore, test results are presented in ranges. Unit operating conditions for all testing was normal full load. For the high velocity tests, gas from both units was combined to achieve a gas velocity greater than 20 ft/sec in the cocurrent section of the module. Table 4 is a general summary of the percent removal of SO_2 at each test condition. A more complete and detailed evaluation of the test results will be available at a later date. The information presented here is only a generalized summary.

Limestone Grind Size

During the low sulfur test program, we operated at both reagent particle size points. A 90% passing 325 mesh was utilized primarily for the zero formic test. A 90% passing 170 mesh was utilized or all other tests. Particle size analysis indicated the target sizes were achieved for all testing.

Table 4

SO_2 Removal Summary

No. of Recycle Pumps	0 Formic Low pH	0 Formic High pH	Low Formic	Design Formic	High Velocity Design Formic
7	88-90%	>94%	>96%	>98%	>98%
6	82-90%	NA	94-96%	97-98%	95-97%
5	68-84%	NA	85-96%	92-97%	93-97%
4	49-76%	55-80%	65-90%	69-95%	91-94%
3	42-61%	50%	58-88%	70-95%	NA
2	30-37%	35%	47-49%	54-56%	NA

Gypsum Dewatering

During the low sulfur test program, gypsum quality was monitored at regular intervals. We did not analyze at every test point. Due to the relative short duration of testing compared to the time to turn over the solids 100% in the module, testing at all points would have been somewhat irrelevant. Laboratory results for the samples that were taken indicated we continued to meet the gypsum quality requirements. Residual moisture was approximately 8%. Residual chlorides were less than 100 ppm. Gypsum purity was greater than 97%.

PROCESS CAPITAL COST AND ECONOMICS

Cost of Demonstration Project

The Cooperative Agreement total project cost of the Milliken Station Clean Coal Technology Demonstration project is $158 million. The total project cost is summarized by project phase in Table 5. The three phases of the project included Pre-award and Design, Construction and Demonstration. Since the project is currently in its demonstration phase, the costs provided in the subsequent analyses, which are based on actual costs for the construction phase, are not expected to change.

The Total Demonstration Project Capital Costs address the total project scope and goals of the demonstration project, in contrast to the scope of the FGD retrofit alone, which represents a portion of the total project scope. These costs are therefore only appropriate if the intent of use is consistent with accomplishing all of the project's demonstration goals, which are identified in the Introduction to this paper.

Table 5. Total Demonstration Project Capital Cost Summary

Project Phase	Original Budget
Phase I (w/Pre-Award)	$ 11,322,048
Phase II	$118,264,240
Phase III	$ 29,021,519
TOTAL PROJECT	$158,607,807

To achieve all of the project's technical objectives, in addition to the FGD system, the $158 million total project cost includes combustion modifications, precipitator modifications, provision for the NOxOUT process, a high-efficiency air heater system and PEOA (Plant Economic Optimization Advisor). Eliminating the non-FGD scope and costs, the resulting actual FGD demonstration technology capital cost is $79 million, or $264 dollars per kilowatt. This adjusted capital cost is summarized by system in Table 6.

The costs in Table 6 represent procurement and installation which occurred during the project's design and construction phase, and therefore reflect mixed year dollars. In addition to the costs of the FGD system, the costs in Table 6 include a new multiple flue stack with FGD by-pass, new ID fans and ductwork, complete limestone receiving and preparation, complete gypsum handling and storage and a separate waste water treating facility.

Table 6
FGD Demonstration Capital Cost Summary

FGD System Titles	Capital $
Limestone Handling and Preparation	5,361,300
Slurry Feed & Recycle	5,736,900
Absorber Module & Auxiliaries	6,570,000
Gypsum Dewatering & Handling	5,337,400
ID Fans & Ductwork	7,464,600
Waste Water Processing System	4,484,100
Other Mechanical Systems	5,356,300
Electrical and I&C	5,886,400
Stack & Flues	2,655,800
FGD Building & Site Work	14,547,500
TOTAL CONSTRUCTION WORK	$63,401,200
Engineering/Construction Management	15,714,900
TOTAL PROJECT COST	$79,116,119

Projected Cost of Commercial Plant

In order to recognize the probable capital cost of this FGD technology in future retrofit applications, without the additional cost burdens associated with a demonstration project, the actual costs of the 2 X 150 MWe demonstration project were adjusted to remove demonstration specific costs. This was accomplished by first reviewing the design basis of nearly thirty process systems. The review identified scope that was necessary for the demonstration, but not representative of a commercial application. This method of modifying the actual costs was extended to equipment, piping systems, instrumentation & controls, foundations and the FGD building. The cost results of this evaluation are summarized in Table 7. Systems which change significantly are Slurry Feed & Recycle, Absorber Module & Auxiliaries, ID Fans & Ductwork, and Stack & Flues.

Table 7
Estimated Cost of 2 x 150 MWe Commercial Plant

FGD System Titles	Capital $
Limestone Handling and Preparation	5,303,000
Slurry Feed & Recycle	4,442,000
Absorber Module & Auxiliaries	6,091,000
Gypsum Dewatering & Handling	4,243,000
ID Fans & Ductwork	7,502,000
Waste Water Processing System	2,363,000
Other Mechanical Systems	4,734,000
Electrical and I&C	5,600,000
Stack & Flues	2,645,000
FGD Building & Site Work	13,416,000
TOTAL CONSTRUCTION WORK	$56,339,000
Engineering/Construction Management	6,766,000
TOTAL PROJECT COST	$63,105,000

For the Commercial Plant, the engineering and construction management cost has been determined as a percent of the capital cost. The resulting cost of $63 million, or $210 dollars per kilowatt, is the anticipated nominal cost for a retrofit FGD serving two 150 megawatt coal fired units.

To establish the expected cost of a nominal FGD system serving a single 300 megawatt unit, rather than two 150 megawatt units, the same basic approach was used, but costs were tailored to the single unit's requirements. The results of this analysis indicate that the nominal cost of a retrofit FGD serving one commercial 300 megawatt coal fired unit is $60 million, or $200 dollars per kilowatt. The results of this evaluation are summarized in Table 8.

Table 8
Estimated Cost of 1 x 300 MWe Commercial Plant

FGD System Titles	Capital $
Limestone Handling and Preparation	5,303,000
Slurry Feed & Recycle	3,668,000
Absorber Module & Auxiliaries	5,552,000
Gypsum Dewatering & Handling	4,243,000
ID Fans & Ductwork	6,323,000
Waste Water Processing System	2,363,000
Other Mechanical Systems	4,734,000
Electrical and I&C	5,600,000
Stack & Flues	2,485,000
FGD Building & Site Work	13,416,000
TOTAL CONSTRUCTION WORK	$53,687,000
Engineering/Construction Management	6,173,000
TOTAL PROJECT COST	$59,860,000

Since many of the cost comparisons that are performed for FGD systems include a nominal 500 megawatt single unit size as a reference point, the 300 MWe system capital cost was used as a basis from which to scale the cost for a hypothetical retrofit of a 500 megawatt unit. The 300 MWe system is a reasonable basis, since the number of modules and system scope do not change, except for the capacities of the systems and equipment. Using the actual cost data for several recent FGD installations which was presented at the 1995 SO_2 Control Symposium1 as a basis, a scaling factor for the retrofit of a FGD system for a 500 MWe unit was calculated.

The results obtained by using the scaling factor calculated from the Symposium data and the size increase of 67 percent indicate that the unit cost of a retrofit FGD system for a 500 megawatt size plant would be $190 dollars per kilowatt.

Applying the unit cost to the 500 MWe unit equates to a capital cost of $95 million, which is nearly 60 percent higher than the cost for the 300 megawatt unit presented in Table 8. Since the resulting scaling factor does not demonstrate a significant cost advantage, costs developed on this basis are considered to be conservative.

Most importantly, the cost is competitive with or superior to other limestone forced oxidation systems. With the expected lowcosts of maintenance associated with the tile lined absorber, operation and maintenance costs for the system will also be very competitive.

Figure 6 uses the capital cost data for several recent FGD retrofit installations presented at the 1995 SO_2 Control Symposium and compares them to the Milliken Station retrofit FGD costs. Figure 6 charts the projects' capital costs ($/kW) relative to plant size. It should be noted that the scope of the Milliken Station FGD retrofit has been identified in detail, while the retrofit scope identified in the symposium data is defined only in terms of the systems' major operating parameters. This leaves room for interpretation, which may result in some inconsistencies in the cost/scope relationship between the individual data points. However, in Figure 6, no attempt has been made to completely normalize scope, or adjust costs accordingly. Some of the possible scope differences could include the following: full scope versus partial scope for sorbent and gypsum or sludge handling systems; sparing of the absorbers; production of marketable by-product; design sulfur removal efficiency; design coal, and the reference year of the reported costs (mixed year dollars over several-year periods).

SUMMARY AND CONCLUSION

Building a major project on a site shared with an operating unit poses unique construction coordination challenges. These challenges were successfully met during construction of the Milliken Clean Coal Technology Demonstration project by adopting an aggressive construction strategy which limited tie-ins of the new process systems to existing unit scheduled maintenance outages. This minimized the loss of station generating time. Also, detailed planning and project execution resulted in the project meeting all critical construction and tie-in dates.

The initial results obtained from the ongoing demonstration of the S-H-U flue gas desulfurization process at Milliken Station appear to be favorable, as SO_2 removal is being achieved at the designed level, and a high quality gypsum by-product is being produced and sold. The planned 36 month demonstration program is intended to demonstrate the effectiveness of the process under various operating conditions, and to establish the system's long term reliability and performance characteristics. The initial test results and the inherent technical and economic benefits of the S-H-U process already enhance its attractiveness as a FGD retrofit alternative in the U.S.

There are both technical and economic advantages to be derived from a formic acid enhanced forced oxidation wet limestone scrubber retrofit similar to that constructed for Units 1 & 2 at the Milliken Station. Most importantly, the S-H-U process is the only developed wet-limestone FGD process designed specifically to employ the combined benefits of low-pH operation, formic acid enhancement, single-loop cocurrent/countercurrent absorption, and in situ forced oxidation. The benefits of operation at lower pH include:

- Reduced limestone requirements
- Limestone does not have to be ground as finely

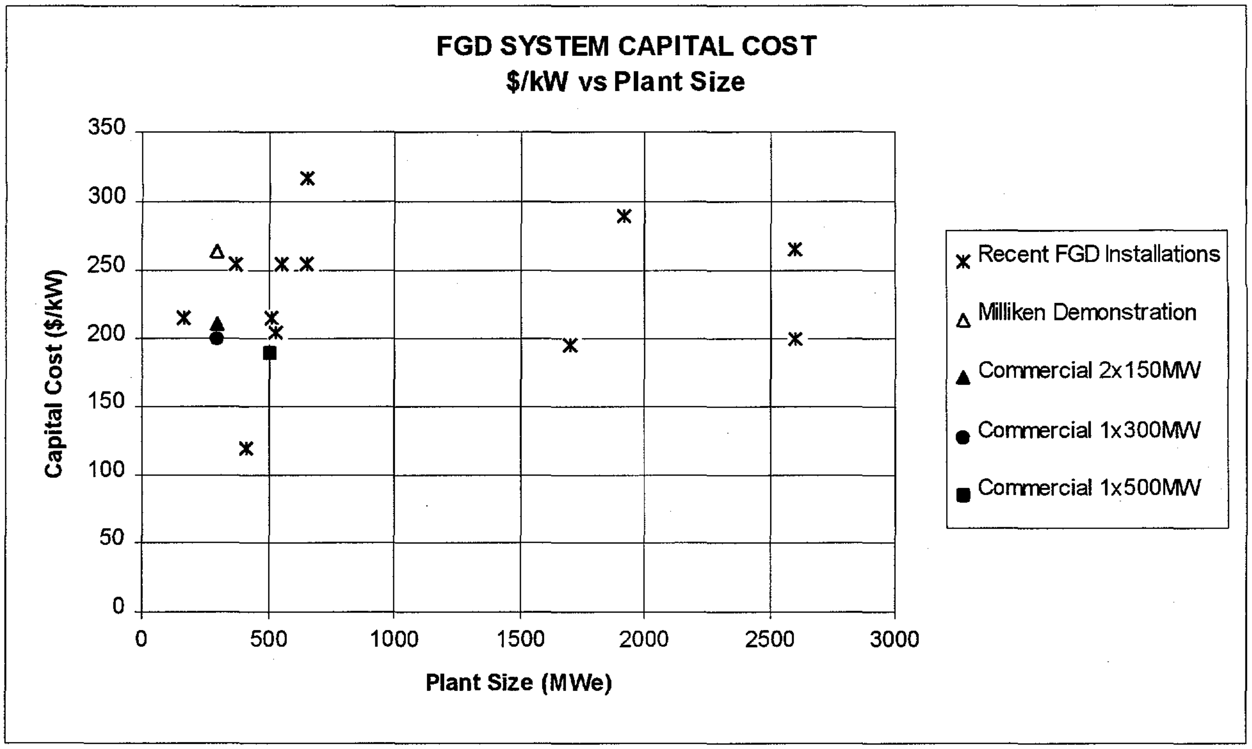


Figure 6

- Less limestone contamination of the gypsum by-product
- More efficient oxidation of the bisulfite reaction product to sulfate
- Reduction in air needed for the oxidation reaction
- More easily dewatered gypsum crystals

Some of the benefits offered by Formic acid buffering are:

- Better SO_2 removal efficiency with limestone
- Lower limestone reagent consumption
- Lower blowdown rate
- Freedom from scaling and plugging
- Higher availability
- Lower maintenance
- Production of wallboard grade by-product
- Improved energy efficiency compared to conventional FGD technologies
- Reduction of Slurry recirculation rates, saving both capital cost and energy, and
- Excellent stability and ease of operation during load changes and transients.

Finally, the process's ability to tolerate higher chloride concentrations reduces the amount of wastewater that must be processed, which in turn results in reduced operating costs.

Aside from the technical and economic advantages of the S-H-U process, other considerations which may determine the applicability of this retrofit limestone forced oxidation FGD system include (1) the availability (or lack) of space to meet the waste disposal requirements of alternative FGD technologies, and (2) sufficient access to construct, operate and maintain an alternative system at the plant site. The S-H-U process can be implemented as a separate facility, or as an integral part of the stack to conserve site space. Because the Milliken Station design consists of a below-stack absorber, this demonstration greatly enhances the viability of the S-H-U wet limestone FGD technology as a retrofit option for other existing plants with space restrictions, as was the case at Milliken Station.

In addition to the impressive operating and construction advantages offered by the formic acid enhanced forced oxidation scrubber, an assessment of the capital costs of retrofits for generating units in the 300 - 500 MW range indicates that a favorable capital cost exists or this system. This makes it highly competitive with other wet limestone FGD technologies. The capital costs for a comparable operating system for plants in the 300 - 500 MW range can be expected to be under $200 per kilowatt. This value compares favorably with capital costs for other wet limestone FGD applications, and the potential exists for even lower costs, depending upon engineering/design optimization.

REFERENCES

1. R.J. Keeth, P.A. Ireland, P. Radcliffe, "Utility Response to Phase I and Phase II Acid Rain Legislation - An Economic Analysis," EPRI/DOE/EPA 1995 SO_2 Control Symposium.