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Project Summary

Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Final Report and Demonstration Test Second Year Results

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Performance of a full-scale flue gas desulfurization unit to demonstrate the Wellman-Lord/Allied Chemical process was evaluated for a period in excess of 2 years. The Wellman-Lord/Allied Chemical process is a regenerable process employing sodium sulfite wet scrubbing, thermal regeneration of the spent scrubber solution, and reduction to elemental sulfur of the recovered SO_2.

Test program results indicate that 89 to 90% of the SO_2 can be readily removed from the flue gas in a long-term dependable manner. Reliability of the coupled absorber and regeneration system for the second year was 61%. For the last 7 months, it was 74%. The major operating limitations were reduction unit problems, but unscheduled outages of the evaporator and the booster blower and start-ups and shutdowns also contributed to down time.

As expected, the energy requirements of the process, primarily for thermal regeneration of the scrubber solution and subsequent recovery of SO_2, were quite large, amounting to 12% of the boiler heat input derived from fuel. Actual annualized operating cost was 14.9 mills/kWh, using 1978 prices for raw materials and utilities. Credits for the sale of byproduct sulfur amounted to only 0.2 mills/kWh.

The reported operation and performance occurred after some modification to the boiler to increase inlet flue gas temperature and after needed improvements to the FGD plant identified during initial operation were implemented. Design limitations affecting overall performance were lack of redundancy, regeneration area capacity of only 80% of full load, underdesign of the purge solids dryer, and limited turndown capability.

This Project Summary was developed by EPA's Industrial Environmental Research Laboratory, Research Triangle Park, NC, to announce key findings of the research project that is fully documented in a separate report of the same title (see Project Report ordering information at back).

Introduction

In 1972, the EPA entered into a cost-shared contract with Northern Indiana Public Service Company (NIPSCO) to design, construct, and operate a flue gas desulfurization (FGD) plant that uses the Wellman-Lord/Allied Chemical (WL/A) FGD process. NIPSCO entered into contracts with Davy Powergas (now Davy McKee) to design and construct the unit and with Allied Chemical (now Allied) to operate the plant. The FGD unit was retrofitted to NIPSCO's Mitchell No. 11 boiler in Gary, IN.

The WL/A process developed by the two design organizations is a regenerable process, based on the recovery of concentrated SO_2 and its subsequent reduction to elemental sulfur. The product is sold to partially offset the process costs. This was the first coal-fired WL application, as well as the first joint WL/A installation.

To ensure that potential users are fully aware of the commercial practicality of the WL/A process, the EPA conducted a test program that included:

- A baseline characterization of the host boiler.
- A performance test required by contract to demonstrate compliance with the performance guarantees.
- Testing for a 2-year period to demonstrate long-term performance and dependability.

This summary highlights results and conclusions of the test program.

Principal objectives of the test program were:

1. Verification of the reduction in pollutants achieved by the WL/A process FGD unit.
2. Validation of the estimated technical and economic performance of the demonstration unit.
3. Assessment of the applicability of the WL/A process to the general population of utility boilers.

The test program was designed to attain these objectives to the maximum extent possible. Emphasis on the various objectives was sometimes redirected due to test program findings and operating difficulties but, in general, the program goals remained unchanged. The test design featured continuous monitoring of SO_2 removal performance. Evaluation of the data was in response to the test objectives and focused on the dependability and economics of SO_2 removal capability while operating the boiler to provide the expected range and variability of the flue gas properties with a single coal type.

Overall Process Design

Mitchell No. 11 is a 115-MW pulverized-coal-fired, balanced-draft boiler with cold-end electrostatic precipitator (ESP) particle control. The boiler was designed to use a coal with a nominal sulfur content slightly above 3% by weight. The FGD unit was designed to accept flue gas at SO_2 concentrations equivalent to this sulfur level in the coal. Flue gas was fed to the FGD plant by the boiler's two induced draft (ID) fans. Before retrofit of the FGD plant, the flue gas went to a stack shared with another boiler. A quick opening damper was installed in the duct to that stack to bypass the FGD plant when not operating and to protect the boiler from damage during upsets. Normally, the FGD plant operated with the bypass damper closed.

The WL/A FGD process removes SO_2 from the flue gas stream by scrubbing with an aqueous sodium sulfite solution and subsequent thermal regeneration to recover the SO_2. The solution is free of solid material. The liberated SO_2 is then reduced to elemental sulfur which is sold. The FGD unit was designed to remove 90% of the SO_2 delivered with the flue gas at flue gas rates equivalent to a boiler load of 92 MW (80% of full boiler load). The processes are proprietary designs of Davy McKee and Allied. Logical separation of the various process steps are:

- SO_2 absorption - Davy.
- SO_2 recovery and scrubber solution regeneration - Davy.
- Purge treatment - Davy.
- SO_2 reduction - Allied.

The following description is based on Davy and Allied non-proprietary design data.

Figure 1 shows the process steps. The FGD plant accepts the total flue gas stream from the discharge of the boiler's ID fans using a booster fan to overcome the flow resistance through a prescrubber and an absorber. The prescrubber is a single-stage orifice contactor designed to remove particulate matter and cool the flue gas before the SO_2 absorption step. A pump recirculates the scrubber water from a sump back to the contactor. In order to control solids buildup in the liquid stream, a purge stream is withdrawn and makeup water is added to the prescrubber to compensate for this loss as well as to humidify the flue gas. The purge stream is sent to the power station's fly ash settling ponds. Particles not removed in the prescrubber are removed with a filter in the spent absorbing solution line leaving the absorber. The wash water from periodic washing of this filter is also discharged to the power station's fly ash settling ponds. These are the only waste streams expected to be discharged from the FGD plant.

The cooled, humidified flue gas leaves the prescrubber and enters the bottom of a three-stage absorber where the gas is contacted with the sulfite solution flowing countercurrent to the gas stream. The solution absorbs the SO_2 and the treated flue gas, saturated at about 54.4°C (130°F), is then discharged to the atmosphere through an integrally mounted stack. Direct-fired reheat of the flue gas with natural gas as fuel was provided but never used because of limited gas supplies.

The absorber is a three-tray column. An absorber demister pad above the top contact stage prevents entrained absorber solution from being exhausted with the treated gas. Each contact stage of the absorber has a separate recirculation system to promote good gas/liquid contact. The absorbing process is conducted at about 54.4°C (130°F), and spent absorbing solution is withdrawn from the bottom contact stage. Regenerated absorbing solution is added to the top contact stage. The process chemistry for absorption is:

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$$

The bisulfite-rich solution is then passed to the evaporator-crystallizer unit. The evaporator-crystallizer is a single-effect forced circulation unit. The heat exchanger employs steam and the clean condensate is discharged for reuse by the host boiler. The heat supplied to the liquor decomposes the sodium bisulfite solution to sodium sulfite (which crystallizes out), SO_2, and water:

$$2NaHSO_3 \xrightarrow{\text{heat}} Na_2SO_3 + SO_2 + H_2O$$

The SO_2 and water vapor are discharged overhead from the evaporator. The wet gas stream is cooled to condense and separate the water vapor from the gas stream, providing a SO_2 feed stream of low humidity for the SO_2 reduction unit. The condensate that was removed is stripped of dissolved SO_2 and is added to the sodium sulfite slurry discharged from the evaporator salt leg. The sodium sulfite and condensate, along with makeup sodium carbonate, are mixed to provide a solids-free sodium sulfite solution suitable for reuse in the absorber.

Oxidation of some of the sodium sulfite during the absorption step to form sodium sulfate is unavoidable:

$$Na_2SO_3 + O_2 \longrightarrow 2Na_2SO_4$$

A portion of the spent absorber solution leaving the absorber is passed through a

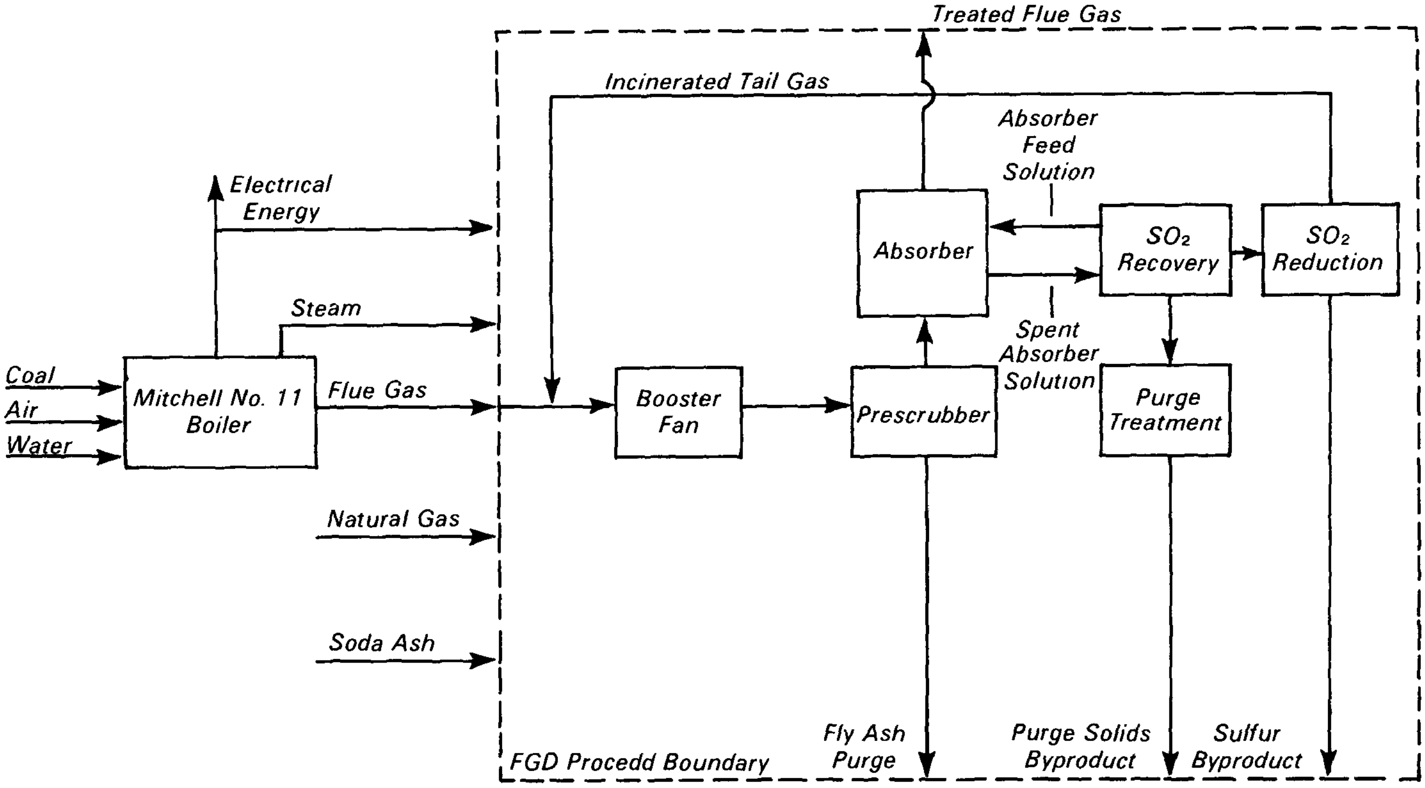

Figure 1. *Block flow diagram of major process steps.*

purge treatment system to separate and dry the Na_2SO_4 for eventual sale. The purge treatment unit was designed to minimize the amount of purge and to produce a salable byproduct of sodium sulfate. Purge requirements are reduced by subjecting part of the absorber product stream to a fractional crystallization process. Sodium sulfate crystals are produced by chilling the solution under very specific conditions. This treatment allows most of the active sodium compounds (sulfite and bisulfite) to remain in the mother liquor which is separated from the crystals by a centrifuge. This mother liquor is then returned to the main process and the crystals are dehydrated in a drying system. The final product is a dry sodium sulfate suitable for marketing. Purge treatment requirements depend on the amount of sodium sulfate and, possibly, sodium thiosulfate formed in the process and on the allowable concentration in the absorbing solution of these inactive components. The amount of sodium sulfate formed may be a function of the amount of excess air in the flue gas.

The SO_2 discharged overhead from the evaporator-crystallizer is reduced to elemental sulfur during the SO_2 reduction step, a two-stage process employing a primary catalytic reaction of SO_2 with natural gas to produce sulfur and some hydrogen sulfide (H_2S) followed by a secondary Claus conversion system in which the H_2S is reacted with residual SO_2 to produce additional sulfur. The primary reaction system consists of two packed-bed regenerative heat exchangers and a catalyst-packed reduction reactor. These vessels and their connecting flues are refractory-lined for protection against high temperatures and corrosive gases.

The regenerative heat exchangers remove the heat from the gases leaving the reactor, and utilize this heat to raise the temperature of the gases entering the reactor. At appropriate intervals, the duties of the two regenerative heat exchangers are alternated; i.e., the packed bed heat exchanger that was heating the entering gases becomes a cooler for the gases leaving the reactor.

Essentially half of the SO_2 in the feed is reduced to elemental sulfur by direct reaction with the reducing gas:

$$2SO_2 + CH_4 \longrightarrow CO_2 + 2H_2O + 2S$$

Simultaneously, most of the remaining SO_2 is converted into H_2S and additional sulfur by a similar reaction:

$$3SO_2 + 2CH_4 \longrightarrow 2CO_2 + 2H_2O + 2H_2S + S$$

The hot gases from the primary reaction system pass through condensers where the sulfur is removed and sent to storage. The gas flows and operating conditions in the primary reaction system are carefully controlled so that the mixture of H_2S and unconverted SO_2 in the product gases from the primary reaction system closely approximates the ideal volumetric ratio (two parts H_2S to one part SO_2) required for the subsequent Claus reaction:

$$2H_2S + SO_2 \longrightarrow 2H_2O + 3S$$

At the same time, maximum utilization of the reducing agent is achieved and the formation of undesirable side-reaction products is minimized.

The Claus conversion system is a two-stage unit. Associated with the Claus unit is an interstage condenser from which elemental sulfur also passes to storage. A final condenser follows the second stage converter to recover the last portion of the elemental sulfur product.

The tail gases from the SO_2 reduction unit pass through an incinerator where natural gas is burned in air to oxidize the remaining H_2S to SO_2. The hot gases from the incinerator are admixed with the untreated flue gases at the booster blower inlet, thus avoiding having to discharge a stream of small volume but containing a relatively high concentration of SO_2.

Design Limitations

The FGD demonstration would have benefited from a more conservative design. Deficient performance was due partially to design related causes. The most significant design decisions affecting performance were:

- The booster blower and the absorber were designed for flue gas flows in excess of boiler full-load flow. The absorber was designed to remove the expected amount of SO_2 at boiler full load. The effect on performance was positive.
- Capacity for recovery of the SO_2 (evaporator, purge treatment, reduction) was based on an expected load factor of 80% of full load. Surge capacity provided by storage tanks for absorber feed solution and absorber spent solution was limited to about 4 hours. The effect was to limit either the amount of SO_2 removed or maximum utilization of the boiler. The latter prevailed because of a policy to operate with a closed bypass and the boiler limited to 80% load factor (92 MW).
- Baseline testing revealed that flue gas flow at a given load exceeded the design flue gas flow significantly. High excess air levels in the flue gas and additional flue gas due to higher-than-expected boiler heat rates were identifiable causes.
- The FGD plant was designed with virtually no redundancy as installed spares.
- Initially, the evaporator circulating pump had a steam-turbine drive. Loss of high pressure steam during boiler shutdowns delayed start-ups because slurry in the evaporator had to be removed and diluted. The deficiency was corrected after 1 year of demonstration with the installation of an electric-motor drive.
- The purge solids dryer was under-designed or a misapplication and this prevented full recovery of sodium sulfate from the purge stream.
- The reduction unit had less turn-down capability than the absorber/evaporator.
- The FGD plant was designed to take flue gas at 149°C (300°F). Initially, actual temperatures were substantially lower due in part to cooling of the flue gas by inleakage air. Some of the baskets were removed from the air heaters to provide flue gas at temperatures of 149°C (300°F) and above.

Test results that were design-related, as opposed to process related causes, have been identified as such in the discussion of test results that follows.

Conclusions

The WL/A demonstration test program consisted of three major test phases:

- Baseline testing.
- Acceptance tests to verify performance guarantees.
- Two-year demonstration test program.

Baseline Testing

Major baseline testing occurred in two time-separated stages. The baseline test was conducted prior to completion of construction of the FGD plant. Flue gas characterizations correlated with boiler operating settings showed that some of the flue gas properties, particularly flow and temperature, differed from those used to design the FGD plant. The differences had a profound effect on the criteria for acceptance testing, and the baseline data were very useful for defining those criteria. The data also were useful for defining the range of testing.

Following completion of the first year of the FGD process demonstration, baseline tests were repeated with several objectives:

- Establish an updated boiler performance baseline for comparison with boiler performance when the FGD plant is operating.
- Obtain an updated characterization of the flue gas leaving the boiler.

The FGD plant was down and completely isolated from the boiler. The collection of additional baseline data was necessary to establish to what extent a substantial rebuild of the boiler since the first baseline test and a change in the type of coal being burned had affected the performance of the boiler. Flue gas volumes, flue gas temperatures, and coal quality were of particular interest because of their impact on FGD operation and performance.

Flue Gas Volume

Flue gas volumes were 27% higher than that used for FGD plant design, 151 m^3/s at 148.9°C (320,000 cfm at 300°F). An excess of inleakage air appears to be the major contributor to the high flue gas volumes, but heat rates that were higher than the new boiler design heat rates probably had some effect. Factors contributing to the higher heat rates were high turbine steam rates and, at low load factors, combustion air in excess of the operating set point. Both the high excess air and the high steam rates add to the volume of flue gas that the FGD plant must treat. The high steam rates would be expected to add to the amount of SO_2 to be removed.

Flue Gas Temperature

Flue gas temperature averaged 148.9°C (300°F) at all load levels during the second baseline test. This was substantially higher than the temperatures measured during the first baseline test. The higher temperatures were due to the removal of part of the heat transfer surfaces of the air preheaters in a deliberate attempt to raise the temperatures above the dew point and thus prevent the scaling and corrosion that was occurring at the FGD booster fan. The temperatures were well above the sulfuric acid dew point.

Coal Quality

The high sulfur (~3%) coal used during FGD operation was burned during the second baseline test and during subsequent demonstration testing. The coal sulfur content was 0.2 to 0.3 wt. % lower than that of the coal burned during the first baseline test. Except for expected differences in variability of the ash and moisture contents, the quality of the two coals was about the same.

Acceptance Test

Process performance guarantees were met or exceeded as confirmed by acceptance testing. The boiler was operated to provide the flue gas flows used to design the FGD plant. Despite flue gas dilution from high excess air levels that was found during baseline testing, the SO_2 feed rates equalled or exceeded the design expectation of 0.6101 kg/s (4842 lb/h).

Specific performance criteria were met or exceeded:

1. SO_2 removal of 90%, 2-hour average, or better was achieved for 261 hours at flue gas flows of 151 m^3/s (320,000 cfm) or higher and was achieved for 84 hours at about 183 m^3/s (388,000 cfm) or higher.
2. Particulate emissions did not exceed 40 ng/J (0.1 lb/10^6 Btu) of boiler heat input at either the lower (design load) or higher load conditions.
3. The consumption of steam, natural gas, and electrical power averaged 76% of the performance guarantee requirements at design load conditions.
4. Soda ash consumption averaged less than 0.069 kg/s (6.6 tons of sodium carbonate per day) which was the limit set in the performance guarantees that was not to be exceeded during the design load period. Sodium carbonate consumption of 0.069 kg/s (6.6 tons/day) is equivalent to 0.152 mol Na/mol S removed from the flue gas at a SO_2 feed rate of 0.6101 kg/s (4842 lb/h) and 90% removal.
5. Sulfur product purity was greater than 99.5% at both design- and high-load conditions.

Two-Year Demonstration Program

The test program, as originally planned, was to monitor performance of the FGD plant for 1 year. The FGD plant achieved only 90 days of accumulated operation during the first year due in part to boiler operating problems. The principal boiler problems that prevented FGD operation were unstable flue gas flows and steam pressures. They were the result of poor coal quality, that exacerbated some problems with the coal feeding equipment, and of boiler feedwater quality problems. Problems were also encountered at the boiler/FGD interface, in particular, booster blower and damper problems. A midyear review resulted in the initiation of a plant improvement program with the goal of correcting the major problems. The program was targeted for completion during a scheduled boiler shutdown which coincided with the end of the 1-year demonstration. The test program was continued for another full year to evaluate the effects of the boiler and FGD plant modifications.

Table 1 lists the major improvement projects. The need for these improvements are identified with the following FGD plant process and design limitations:

- Steam and flue gas fluctuations limited or prevented reliable operation.
- Flue gas temperatures below the acid dew point caused unbalancing of the booster blower and (ultimately) severe corrosion/erosion damage to the fan blades.
- The guillotine-type damper installed to isolate the FGD plant from the boiler during shutdowns became inoperable, after binding from the accumulation of aggregated fly ash along the tracks.
- Energy supply to drive the evaporator circulating pump must not be interrupted. Originally, this drive was a steam turbine operating on high pressure steam supplied by the boiler. During a boiler shutdown and with no evaporator circulation, slurry in the evaporator had to be drained and diluted at considerable penalty in additional startup time required.
- Sulfur condenser leaks were a recurring problem throughout the 2-year demonstration.

Test program results and operating experience during the second year showed a substantial improvement in FGD performance; since boiler utilization was high, the results more nearly duplicated those expected during commercial application. Flue gas characteristics were essentially unchanged from those of the second baseline test. The conclusions that follow are based on second-year results.

Dependability

Reliability of the FGD unit (hours operated/hours called upon to operate) was 61%. However, FGD plant utilization was consistent for the last 7 months: reliability averaged 74% for that period. Only full operation was counted as hours operated, full operation being integrated operation of the absorber/evaporator loop with the reduction unit when the flue gas bypass was closed. The purge treatment unit

Table 1. *Plant Improvement Projects*

Item	*Date completed*	*Action*
Coal supply	*Completed June 78*	*Provide an uninterrupted supply of Captain coal for Mitchell No. 11 use.*
Air heater	*During September 78 shutdown*	*Remove part of baskets which provide heat storage, to raise inlet duct temperature.*
Duct insulation	*After September 78 shutdown*	*Insulate duct before and after booster blower.*
Blanks	*During September 78 shutdown*	*Provide way to install blanks rapidly at inlet of booster fan as an alternative to the isolation damper.*
Booster blower	*Not completed*	*Install a sparger pipe in the booster blower to periodically steam clean blades while running.*
Evaporator pump	*During September 78 shutdown*	*Install an electric motor as an alternative to steam turbine drive.*
Absorber	*During September 78 shutdown*	*Recoat and repair leaks.*
Booster blower turbine	*After September 78 shutdown*	*Provide enclosure to protect against SO_2 and weak acid attack.*
Sulfur condenser	*During September 78 shutdown*	*Plug leaking tubes.*

Note: Dates indicated were during first year of demonstration test program, September 1977-September 1978.

may or may not have been operating. The reliability record was established with virtually no redundancy built into the FGD unit. Also, the evaporator was designed for the equivalent of only 80% of full boiler load. With limited surge capacity in the regeneration loop, the FGD plant was unable to effect complete SO_2 recovery during evaporator or reduction unit shutdowns. The FGD plant was utilized (hours operated/hours in period) 56% of the time. The downtime was due to:

- FGD repair - 33%.
- Start-up and shutdown - 5%.
- Boiler down - 6%.

The SO_2 absorber was essentially trouble free. The reduction unit required the most downtime for repairs, followed by the evaporator circulating pump and the booster blower (Table 2). Called upon time is defined as the time the boiler operated to provide flue gas and utilities and the time other feed streams were available within the specific design criteria of the FGD plant. These include:

- Flue gas at rates not less than 46 MW equivalent.
- Stable steam pressures within an operable range around a design pressure of 3,790 kPa (550 psig).
- Electricity.
- Natural gas.
- Soda ash.
- Boiler stable within limits of greater than 46 gross MWe, and coal sulfur content greater than 2.8 and less than 3.5 wt. %.

SO_2 Removal

The process was controlled to remove 89% of the inlet SO_2. Thirty-day average removal efficiencies varied from 88% to 93% (Figure 2). The 24-hour and 1-hour average data for removal efficiency were:

SO_2 removal	Percent of time operated: 24-h average	Percent of time operated: 1-h average
90% and greater	60	52
89% and greater	84	78
85% and greater	97	97

On a boiler heat input basis, SO_2 emissions were controlled in the range of 110 to 400 ng/J (0.25 to 0.94 lb/10^6 Btu).

SO_2 removal was attained at electrical generating outputs in the range of 53 to 85 MW of the 115 MW boiler. The lower limit was set by the limiting turndown capability of the reduction unit. The upper limit was set by the 80% capacity limitation of the evaporator, as designed. Because a substantial amount of energy (primarily as boiler main steam) was consumed by the FGD plant, the generating potential of the boiler was actually about 95 MW at the FGD maximum capacity limit of 85 MW. "Generating potential" refers to the gross megawatts that the boiler is capable of generating during FGD operation but cannot attain due to the boiler main steam consumed by the FGD plant.

Table 2. Reasons for Interruption of Operations

Equipment or reasons	*Days of interruption*	*% of called upon time*
Reduction unit	*59*	*17*
Evaporator circulating pump	*28*	*8*
Booster blower	*18*	*5*
Start-up and shutdown	*17*	*5*
Other equipment, including absorber	*10*	*3*
Evaporator	*8*	*2*

Confirmation of Design Limitations

Design limitations were identified in the Introduction. The effect on FGD plant performance was:

- Performance at maximum design flue gas flow rates and SO_2 feed rates could not be adequately tested because of the design limitations of the regeneration system. Flue gas volumes up to 95% of full boiler load were treated successfully during the acceptance test. During the demonstration test program, the maximum SO_2 feed rate that was sustainable for 24 hours was 0.750 kg/s (5950 lb/h), exceeding design expectations.
- Capacity of the recovery unit met design expectations. The recovery unit successfully processed 90% of an SO_2 feed rate of 0.588 kg/s (4670 lb/h) at a flue gas flow rate of 152 m^3/s (323,000 acfm). The boiler generating potential was 89 MW and actual gross electrical generating output was 80 MW. This output was sustained for 65 hours during maximum load tests.
- Baseline flue gas flows were confirmed. At a generating poten-

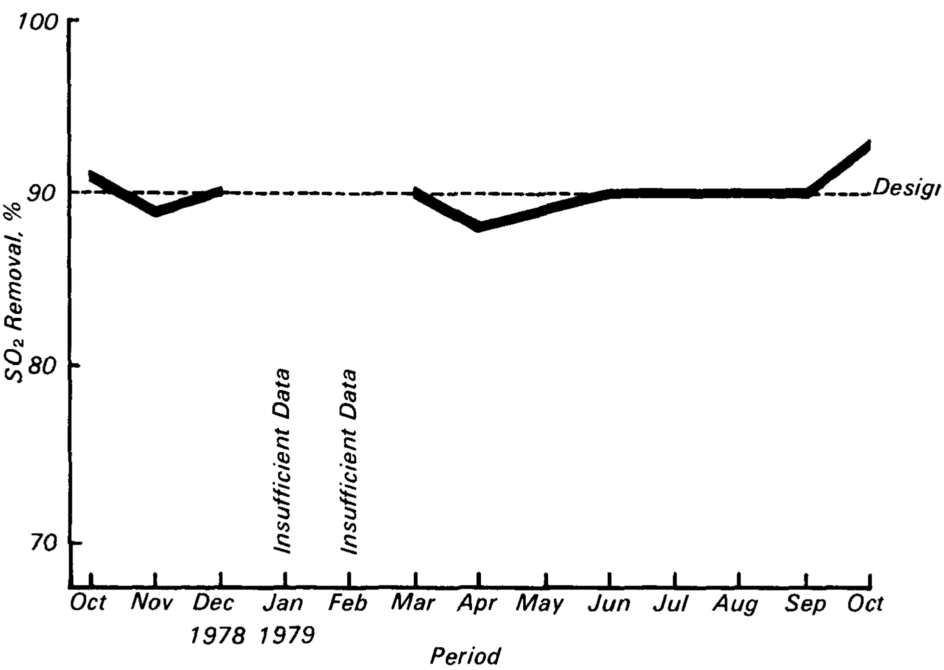

Figure 2. Thirty-day average SO_2 removal.

tial of 92 MW, the design point, the flue gas flow rate was 170 m^3/s (360,000 acfm). This was 13% higher than design expectations. Oxygen in the flue gas averaged 8.0% by volume, compared to an expected oxygen level of 5.6%.

- Installation of an electric drive for the evaporator circulating pump reduced start-up time significantly.
- Tests of absorber turndown, with the reduction unit not operating, established a minimum throughput equivalent to a generating potential of 50 MW. The limiting factor was a flue gas flow rate of 100 m^3/s (220,000 acfm), limited by the minimum governor setting of the steam turbine drive of the booster blower. The minimum sustainable load at full operation of the FGD plant, with the reduction unit operating, was a 62 MW generating potential. Ninety percent SO_2 removal was achieved for 4 days at an average SO_2 feed rate of 0.426 kg/s (3380 lb/h).

Process Economics

The projected annual operating cost in 1978 dollars was 13.2 mills/kWh. Actual annualized cost for the second year of the demonstration, using 1978 prices for raw materials and utilities, was 14.9 mills/kWh. Byproduct sulfur production averaged 1.81 kg/s (17.3 ton/day). Credits for the sale of the sulfur amounted to only 0.2 mills/kWh. Actual annualized cost was based on the 82.2 MW of generator output that was possible at the FGD design capacity of 92 MW, assuming 351 days of boiler operation per year. On this basis, the actual boiler capacity in total kilowatt hours was 92% of the projected capacity. Annualized costs are quite sensitive to lower-than-projected capacities because fixed costs, about 50% of the annual costs, and labor costs continue to accrue whether or not the boiler or the FGD plant is operating at full capacity.

Energy and Raw Material Consumption

A significant amount of steam produced by the boiler was consumed by the FGD plant, used primarily by the evaporator to recover SO_2 and regenerate the scrubber solution (Figure 3). Electrical power consumption amounted to about 1 MW after the evaporator circulating pump had been converted from steam-turbine to electrical drive for improved operability. Actual average steam consumption was 105% of design expectations. The energy equivalent of this steam was 11% of the boiler input energy derived from fuel. Since the average generating output of the boiler was 77 MW, the equivalent loss in electrical generating capacity amounted to an 8% derating of the boiler from a nameplate capacity of 115 MW. Including 1 MW of electricity consumed, the total energy requirement was 12% of the boiler heat input derived from fuel at an average boiler load of 77 MW.

Soda ash was used as makeup sodium carbonate for the scrubbing process. Makeup is made necessary by the buildup of inactive constituents in the absorber/evaporator loop, such as sulfate and thiosulfate, that must be purged. Any loss from the system due to leaks also would require soda ash makeup. High soda ash consumption during the first demonstration year was due to leaks at the bottom collector tray of the absorber that were repaired before commencing the second demonstration year. Average daily consumption of soda ash for the last 7 months of operation was 0.091 kg/s (8.7 tons/day), using the total operating days of the absorber/evaporator as the time base. Soda ash consumption as a function of SO_2 removed was 0.217 mol Na/mol S removed. The performance guarantee for acceptance was 0.069 kg/s (6.6 tons/day) at the design levels for flue gas flow and inlet SO_2. Soda ash consumption of 0.069 kg/s (6.6 tons/day) is equivalent to 0.152 mol Na/mol S removed at a design feed rate of SO_2 of 0.6101 kg/s (4842 lb/h) and 90% removal.

Natural gas was used as the reductant for converting the SO_2 to elemental sulfur. It also was the fuel used to incinerate the tail gas emitted from the reduction process. The tail gas was returned to the inlet of the absorber after incineration. It was necessary to continue incinerator operation during shutdowns to destroy the reduced sulfur forms that desorb from the reduction unit refractory materials. Thus, there was a corresponding improvement in unit consumption of natural gas with improvement in reliability. About 0.2 m^3 (7 ft^3) of natural gas was consumed per pound of sulfur produced which was in accordance with the design expectations. During operation, the incinerator consumed 7.5% of the natural gas. In contrast, the incinerator consumed over 12% of the gas overall because it continued to operate during shutdowns, demonstrating the

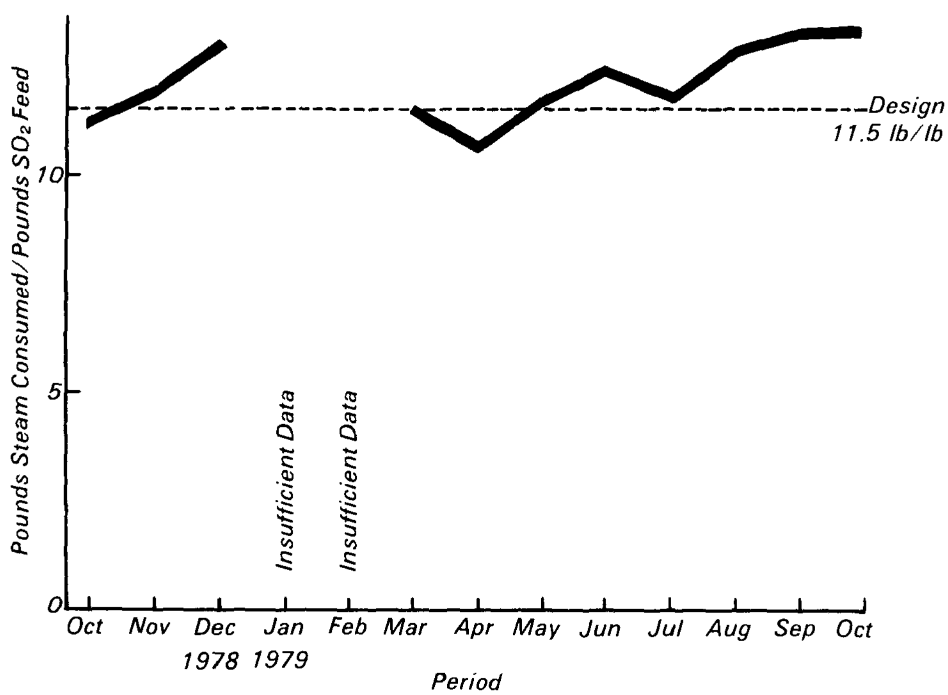

Figure 3. *Comparison of actual steam consumed as a function of SO_2 feed rate with design value.*

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result of a 61% FGD plant reliability factor

Purge Treatment Considerations

The purge unit, as initially designed, was to have treated a small purge stream removed from the regeneration loop, to separate sodium sulfate from most of the sulfite/bisulfite components, and to dry the sodium sulfate to produce a marketable product. The "wet" end of this purge treatment system performed satisfactorily. The limitation of the purge dryer already has been mentioned.

The amount of purge to be treated is a function of the formation of sulfate and possibly thiosulfate. Attempts to determine the amount of sulfate formation during absorption were frustrated by an inability to obtain correct flow measurements and uncertainties about the specific water balance across the absorber. However, the data seem to indicate that sulfate formation is a function of the oxygen concentration in the flue gas. Thus, higher than design purge rates might have been due to the high excess air levels in the flue gas. The average purge rate* for the last 7 months of operation has been estimated to be 18.2% to 24.8%, substantially higher than expected. The estimate was determined from soda ash consumption and the calculated amount of SO_2 removed. A purge rate of about 10% was the value indicated during the design phase of the project. In summary, the information seems to indicate actual purge rates much higher than design.

*Purge rate was determined as the ratio of moles sodium consumed to moles SO_2 removed from flue gas, expressed as a percentage

Purge rates of this magnitude put a further load on the purge solids dryer. Dryer tests performed by Davy McKee determined that the dryer did not have the needed capacity, even at design rates. The maximum dryer capacity achieved during the test, approximately 66% of the design heat duty, could not be sustained because of a buildup of solids at the discharge end of the dryer. Maximum capacity that could be sustained without this buildup was only 45-50% of design.

Davy McKee is investigating the use of an antioxidant that shows promise for reducing the oxidation rate and, thus, the amount of sodium values that must be purged as sodium sulfate.

Recommendations

Overall performance of the FGD demonstration unit was affected significantly by design limitations that were known but not eliminated for various reasons. Typically, capital cost saving funding limitations, and need for furth development are the incentives for le conservative design. For installatior demonstrating new FGD technolog design criteria should be established the start of the program that focus c the advantages and limitations of th process rather than having to repo poor performance solely because design limitations. The WL/A demor stration plant was design-limited by lac of regeneration capacity and by almo complete lack of installed spare Improved performance would be e pected with a more conservative desig It is recommended that full or exces capacity and redundant equipment b designed into future demonstrations t the maximum extent possible. Th demonstration test and evaluatio would have to indicate the degree c overdesign and associated costs, if an so that installed costs relative t performance are demonstrated.

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The complete report, entitled "Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Final Report and Demonstration Test Second Year Results," (Order No. PB 81-246 316; Cost: $29.00, subject to change) will be available only from:

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