

## A NOVEL, ENERGY-EFFICIENT SO<sub>3</sub> FLUE GAS CONDITIONING PROCESS

William G. Hankins  
and  
Ray E. George  
The Chemithon Corporation  
5430 West Marginal Way SW  
Seattle, WA 98106

### Abstract

For the past 20 years, sulfur trioxide (SO<sub>3</sub>) flue gas conditioning (FGC) systems have been installed primarily to enhance the performance of precipitators that failed to meet design expectations while burning low sulfur coals.

More recently, FGC has been viewed as part of a cost effective strategy to comply with the new Clean Air Act (CAA) of 1990. The law requires annual sulfur dioxide (SO<sub>2</sub>) emission rates to be reduced below 1980 levels by 10 million tons. Sources which were not affected by the original CAA will have to comply with the new CAA. As a result of these legal changes, pressure is being applied for power plant executives to develop compliance strategies.

FGC is now being used as part of an overall strategy to reduce SO<sub>2</sub> emissions by switching to low sulfur fuels. The capital cost is a fraction of alternative methods, such as scrubbing. However, operating costs for an FGC system can be significant. For a large pulverized coal fired boiler, typical operating costs may be about \$175,000 annually. Systems operating with low burner SO<sub>2</sub> concentrations would be even more expensive to run. To optimize the operational cost effectiveness of FGC hardware, The Chemithon Corporation developed some new approaches to the system process. The cost savings of the

new system design were recently demonstrated on the two new FGC systems installed at Georgia Power Bowen Units 3 and 4.

The projected operating costs were compared with actual operating conditions at Plant Bowen. These are 880 MW boilers, each operating with the improved FGC design (patent pending) Chemithon developed. Measured and observed data were in close agreement with values projected by computer modeling. The higher SO<sub>3</sub> concentrations and pressures allowed by the new design also reduced the installation costs. This was primarily due to the smaller pipe used for the gas delivery and injection systems.

## Introduction

The history of flue gas conditioning dates back almost as far as the first electrostatic precipitator (ESP). As early as 1912 it was discovered that increasing levels of SO<sub>3</sub> in smelter converter gases increased the collection efficiency of the ESP. Experimental work demonstrated that many non-conductive dusts and fumes could be made collectable by adding SO<sub>3</sub> and/or moisture to the gas stream ahead of the ESP<sup>1</sup>. Since that time, many other substances have been used to condition flue gases. These include: ammonia, triethyl amine, and various proprietary chemicals. SO<sub>3</sub> is still the predominate treatment for high ash resistivity problems.

For pulverized coal fired furnaces, reduction in fuel sulfur below design levels frequently causes difficulties for ESPs. The performance of electrostatic precipitators is sensitive to the physical properties of flue gases and fly ash particles, particularly ash resistivity. High fly ash resistivities, above about  $1 \times 10^{10}$  ohm-cm, may result in decreased collection efficiencies. This problem can generally be reversed by injecting SO<sub>3</sub> upstream from the ESP. The SO<sub>3</sub> condenses on the particulate, improving its ability to accept and release an electrical charge.

The Clean Air Act of 1990 requires annual SO<sub>2</sub> emission rates to be reduced below 1980 levels by 10 million tons. Sources which were not affected by the original CAA will have to comply with the new CAA. Power plant executives must develop compliance strategies that are effective and affordable. Fuel switching, in combination with FGC, has been shown to be a cost effective means to reduce SO<sub>2</sub> emissions to levels mandated by the law. The capital cost is a fraction of installing and operating scrubbing systems. However, operating costs for an FGC system can still be significant.

To optimize the performance of SO<sub>3</sub> conditioning systems while reducing operating costs, The Chemithon Corporation sought to improve the hardware design

of the process gas system. The improvements developed and tested are discussed below.

The two units recently installed at Georgia Power Plant Bowen Units 3 and 4 incorporated these cost saving design improvements. These improvements are discussed and compared to prior art systems in this paper.

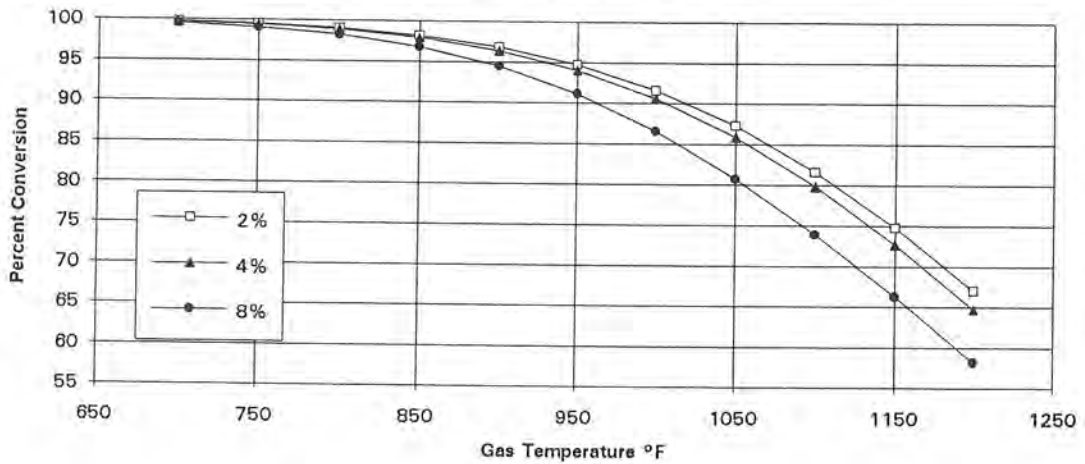
## Process Description

In sulfur burning  $\text{SO}_3$  FGC systems, molten sulfur is delivered to a steam heated storage tank. Molten sulfur is pumped from the storage tank to the sulfur burner where it mixes with air and burns to form  $\text{SO}_2$ . The hot  $\text{SO}_2$ /gas mixture enters a catalytic converter where the  $\text{SO}_2$  oxidizes to form  $\text{SO}_3$ . The hot  $\text{SO}_3$ /gas mixture is then conveyed to an array of injectors located in the ductwork ahead of the electrostatic precipitator. Typically,  $\text{SO}_3$  is injected at rates to achieve 5 to 25 ppm in the flue gas. For an 880 MW boiler an injection rate of 25 ppm is equivalent to an  $\text{SO}_3$  production rate of 710 lb/hr or about 300 lb/hr of sulfur.

To achieve conversion of the  $\text{SO}_2$  to  $\text{SO}_3$  within the catalytic converter, the temperatures entering and exiting the converter must be within a specific range. Typical catalysts convert  $\text{SO}_2$  to  $\text{SO}_3$  in the range of 750 to 1100°F. The conversion of  $\text{SO}_2$  to  $\text{SO}_3$  within the converter is exothermic. As the temperature of the reaction approaches 1100°F, the chemical equilibrium tends to favor a reverse reaction back to  $\text{SO}_2$ . Therefore it is important to initiate the reaction at the lowest practical temperature for good conversion. Figure 1 illustrates the temperature dependent conversion equilibrium for various gas concentrations. The closeness to which the curve can be approached is influenced by a variety of parameters, such as: gas velocity, gas distribution in the bed, and catalytic activity<sup>2</sup>.

As seen in the figure, the effect of temperature on conversion efficiency attenuates as the temperature is reduced. Below about 900°F, the effect on efficiency due to varying the  $\text{SO}_3$  concentration becomes minimal. Below about 800°F, the conversion efficiency is nearly independent of both temperature and gas strength.

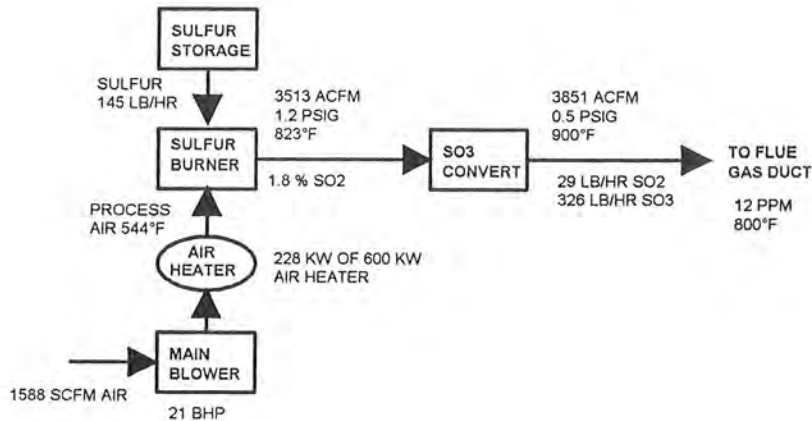
Figure 1  
Equilibrium Conversion for SO<sub>2</sub> Oxidation



Experience has shown that the optimum temperatures for the SO<sub>2</sub> mixture to enter the converter are between 780°F and 850°F, depending on the characteristics of the system. To ensure that the SO<sub>2</sub> mixture enters the converter in this temperature range, most systems rely on a process air heater to raise the temperature of the air entering the burner. This supplements the heat provided by the sulfur combustion. The process air heater is also used to heat the air entering the burner to at least 650°F to light the sulfur for system startup.

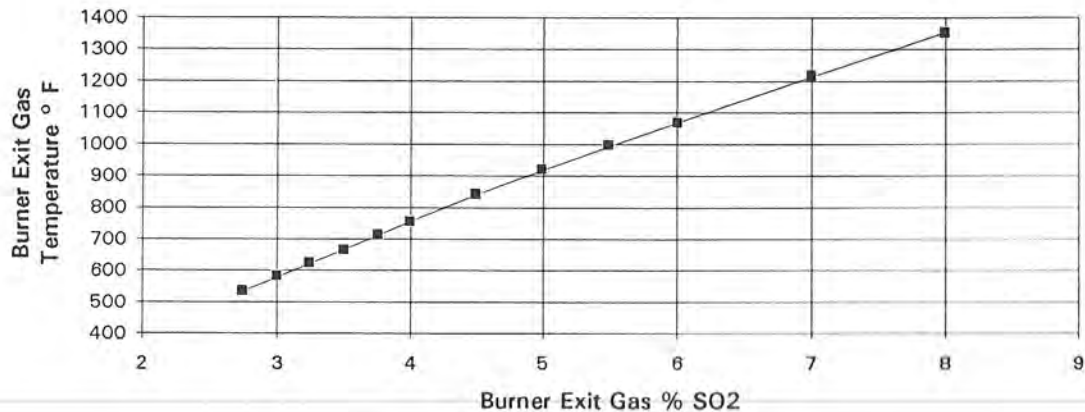
Figure 2 is a simplified diagram of a typical sulfur burning FGC system indicating approximate flow rates and temperatures of the process at about 45% design capacity of 710 lb/hr SO<sub>3</sub>. The system utilizes a fixed speed blower or fan to provide process air. The process air fan is generally sized to provide a burner exit SO<sub>2</sub> gas strength between 2 and 4% at the maximum design production rate. With a 4% SO<sub>2</sub> gas strength, generally little or no process air heat is required at the full production rate to achieve a burner exit temperature of about 820°F, which is in the desired converter inlet temperature range. Therefore, as the sulfur burn rate is reduced, the process air heater exit temperature must be increased to maintain this temperature. Each pound of sulfur burned yields approximately 4000 BTUs.

**Figure 2**  
**Typical Sulfur Burning**  
**Flue Gas Conditioning System**



Another concern in designing the system is to maintain the SO<sub>3</sub> gas temperatures from the converter to the injectors above the acid dew point at all times. Figure 3 shows predicted burner exit temperature vs. SO<sub>2</sub> gas strength without the addition of process air heat.

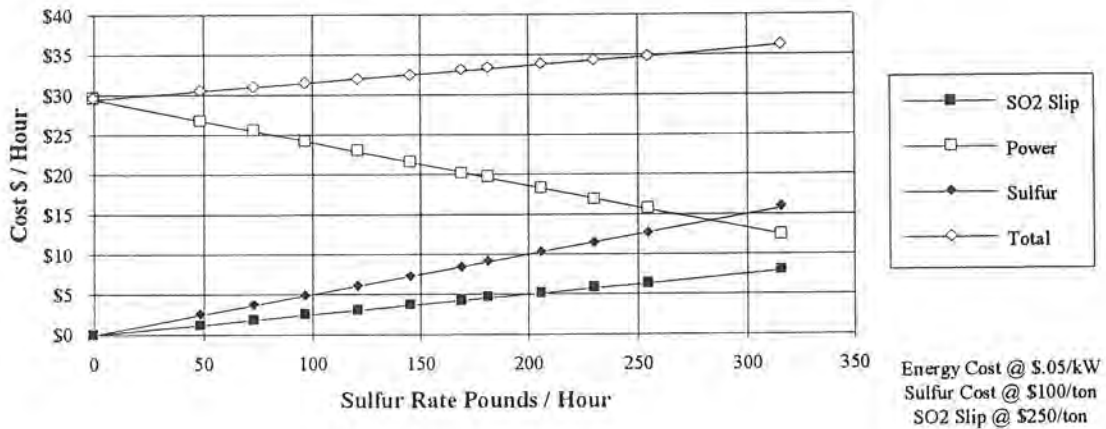
**Figure 3**  
**Sulfur Burner Exit Temperatures**  
**vs. SO<sub>2</sub> Gas Strength**



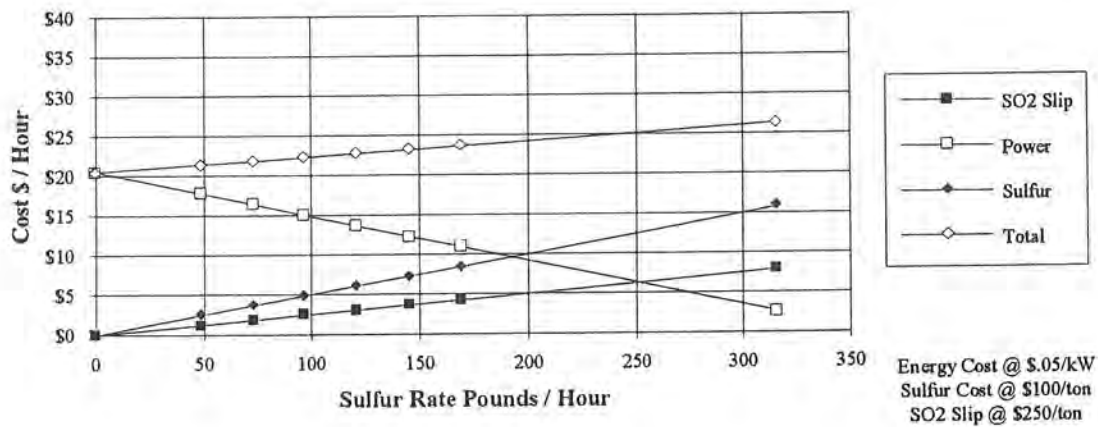
Another consideration of design gas concentration is the effect on converter efficiency. With a gas strength of 4% SO<sub>2</sub>, the converter outlet temperature from a single stage converter will be approximately 1010°F. As seen in Figure 1, the maximum converter efficiency for this condition would be approximately 90%. One might conclude from this graph that designing a gas conditioning system for a low SO<sub>2</sub> burner concentration can be an easy way to achieve high gas conversion. The converter inlet temperatures can be controlled by the process air heater for variations in sulfur rate and the exotherm across the converter stage would be lower, yielding better conversion. In addition, there would be a large surplus of oxygen in the converter.

However, there are significant disadvantages with this approach. It would require increasing the size of the equipment due to the large gas volume, greatly increasing the installed costs. The high volume requirements for heated air would substantially raise operating costs. This is illustrated in Figures 4 and 5. The figures show predicted air heater and gas generator total power costs vs. production rate for systems designed for 2.75% and 4% SO<sub>2</sub> at the maximum rate of 300 lb/hr of sulfur.

**Figure 4**  
**Predicted Gas Generator Operating Costs**  
**for 2.75% SO<sub>2</sub> Concentration**



**Figure 5**  
**Predicted Gas Generator Operating Costs**  
**for 4.0% SO<sub>2</sub> Concentration**



These problems warranted the development of a new design, such as that installed at Georgia Power Bowen Units 3 and 4. The new design offers significant improvements to reduce operating and installed costs.

***Description of FGC Equipment and Process Improvements at Georgia Power Bowen Units 3 & 4***

Two units of the new design were installed at the Georgia Power Co. Plant Bowen on Unit numbers 3 and 4. Each FGC unit was designed to burn 300 lb/hr sulfur, and to provide a 25 ppm SO<sub>3</sub> treat rate for an 880 MW pulverized coal fired boiler. The volumetric flow rates for the flue gas streams, at the points of FGC gas injection, were about 3.2(10)<sup>6</sup> ACFM at 320°F. The SO<sub>3</sub> was injected upstream of Buell electrostatic precipitators.

The improvements in system efficiency and operating costs were achieved by adding SO<sub>2</sub> coolers and high pressure air delivery systems. These devices also affect the operation of the burners, catalytic converters, and gas delivery systems.

The SO<sub>2</sub> cooler was a concentric tube heat exchanger. The SO<sub>2</sub> combustion gases flowed through the inner tube and the cold air flowed through the annulus between the inner and outer tubes. A counter-flow arrangement was used, where the cold air entered at the same end which the cooled gas mixture exited.

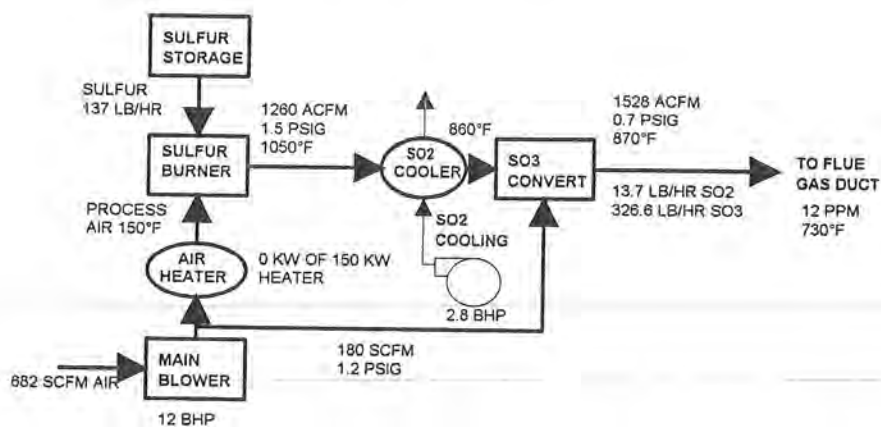
The flow of air through the SO<sub>2</sub> cooler was driven by a constant speed centrifugal fan. The air flow rate was controlled by varying the position of an in-line damper. The damper setting was determined according to the temperature of the SO<sub>2</sub>/gas mixture that entered the catalytic converter.

Process air was supplied by a single, lobe-type, positive displacement (PD) blower. An important feature of the positive displacement design was that flow rates were not influenced by changes in downstream pressure, unlike centrifugal blowers. A variable speed electric motor was used for the blower drive mechanism. This device allowed precise control of the air flow rates and high system pressures; the blower compressed filtered air to about 6 psig (at design flows) for process needs. The variable frequency drive (VFD) minimized power requirements for moving process air at reduced loads.

The two SO<sub>3</sub> FGC systems installed at Bowen 3 and 4 included improved hardware to allow the units to operate with 6 to 8% SO<sub>2</sub> gas strengths at the burner outlet. These concentrations were maintained over most of the operating range. To accomplish this, the process air flow was varied with the sulfur flow to maintain a consistent SO<sub>2</sub> gas concentration exiting the burner.

Due to the high gas strength at production rates above about 6 ppm SO<sub>3</sub>, the gases exiting the sulfur burner had to be cooled prior to entering the converter. Air flow to the SO<sub>2</sub> cooler was varied by adjusting the control valve to achieve the desired converter inlet temperature. At maximum system production the burner exit temperature was reduced from approximately 1200°F to about 820°F. High burner temperature ensures complete combustion of sulfur, minimizing the potential for sublimed sulfur carryover to the converter. Figure 6 illustrates the process and flow information about 45% capacity.

**Figure 6**  
Improved Sulfur Burning  
Flue Gas Conditioning System





A two-stage catalytic converter was used in the design. The converter was sized to provide a minimum of 95% conversion efficiency. Process air was used to cool the gases exiting the first stage and to supply additional oxygen for the reaction in the second stage. The burner air flow and the converter dilution air were controlled in proportion to the sulfur flow. This allowed the system to quickly reach equilibrium temperature and avoid problems such as process air heater temperature oscillation. The temperature into the converter, which was the only critical temperature of the system, was controlled directly to the desired temperature by the SO<sub>2</sub> cooler.

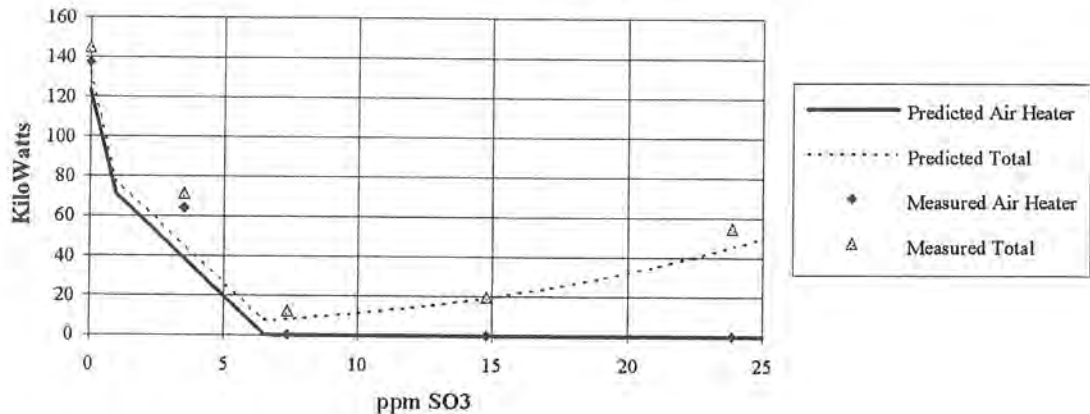
## Test Results

The test results and operating data closely match the computer modeling and design parameters. The results of converter efficiency testing performed by Southern Research Institute at Plant Bowen on the FGC system showed measured conversion of SO<sub>2</sub> to SO<sub>3</sub> at 25 ppm SO<sub>3</sub> injection rate of: 94.1%, 95.6%, and 96.4%.

Gas generator power measurements were made on Unit 3 at several operating modes from standby to 275 lb/hr of sulfur flow. In standby, the equipment was kept hot without burning sulfur. Figure 7 shows the results of the power measurements compared to the computer modeling. The measured air heater power at sulfur rates below about 85 lb/hr sulfur (approximately 7 ppm SO<sub>3</sub>) was slightly higher than predicted by the model. This was believed to be caused by a difference in the minimum burner air flow volume at low sulfur rates.

The temperature loss through the hot gas piping was found to be within 20°F of the temperatures predicted by the computer model at all modes of operation. The length of pipe between the converter outlet and the point of injection was about 150 ft. In the standby mode of operation the injector temperatures averaged 610°F. At full system capacity the injector temperatures measured an average of about 760°F.

**Figure 7**  
**Bowen Unit 3 Gas Generator Skid**  
**Measured Power Consumption vs. Modeled**

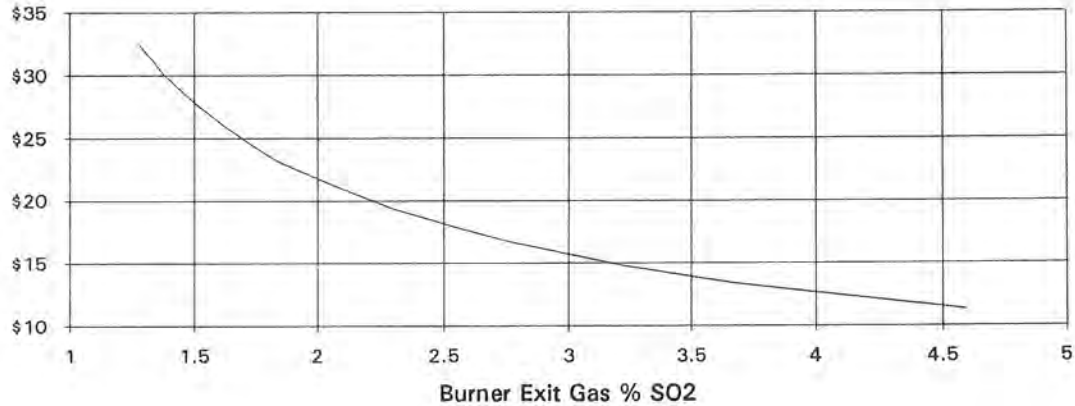


These same design principles were applied to a unit recently installed at Dairyland Power Genoa Unit 3. This unit was designed for a maximum capacity of 100 lb/hr of sulfur. The testing, performed by Clean Air Engineering, indicated conversion of more than 96% at 25 ppm SO<sub>3</sub> generation rate (98 lb/hr sulfur). The converter inlet SO<sub>2</sub> concentrations measured an average of 8.4% SO<sub>2</sub> by volume.

Based on the computer model and data collected from Bowen and Genoa stations, operating costs can be estimated based in part on burner gas concentration. Prior art systems which use fixed air flow into the burner and no cooling of the burner exit gases are designed so that the burner SO<sub>2</sub> concentration at maximum capacity is less than about 4%. For a system designed to burn approximately 300 lb/hr of sulfur producing 25 ppm the process air heater would consume little or no energy at full capacity.

At about half capacity (12 ppm) the system would have a gas strength of about 2% and would require about 250 kW of process air heater. The cost of this design is illustrated in Figure 8. It can be observed from this curve that systems which do not reduce burner air flow with decreasing sulfur rate can be very expensive if they operate at substantially less than the design maximum capacity.

**Figure 8**  
**Hourly Operating Cost at 12 ppm vs. SO<sub>2</sub> Gas Strength**  
**for 300 lbs/hr Max Rate System**



Based on the measured power consumption at Bowen Unit 3, Figure 9 shows estimated operating cost vs. ppm injection rate. Included on this figure are the assumptions of the sulfur cost, energy cost, and SO<sub>2</sub> slip. The SO<sub>2</sub> slip assumes that SO<sub>2</sub> not converted to SO<sub>3</sub> by the system and ultimately released up the stack has a cost. At Plant Bowen, a 25 ppm SO<sub>3</sub> injection rate at full boiler load is equivalent to about 275 lb/hr of sulfur. The operating costs are lowest at about 7 ppm. Experience shows that required SO<sub>3</sub> feed rates are frequently less than 15 ppm at full boiler load for many installations and coals.

**Figure 9**  
**Georgia Power Bowen Unit 3**  
**Expected Gas Generator Operating Costs**

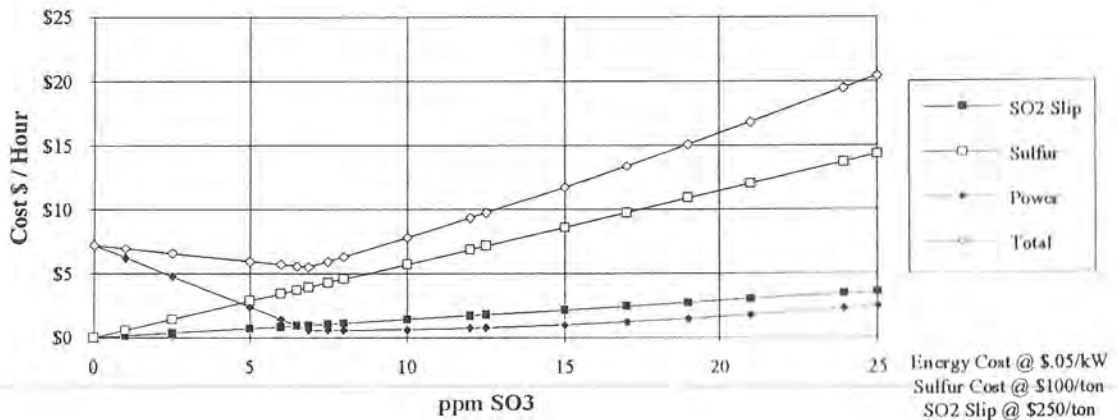
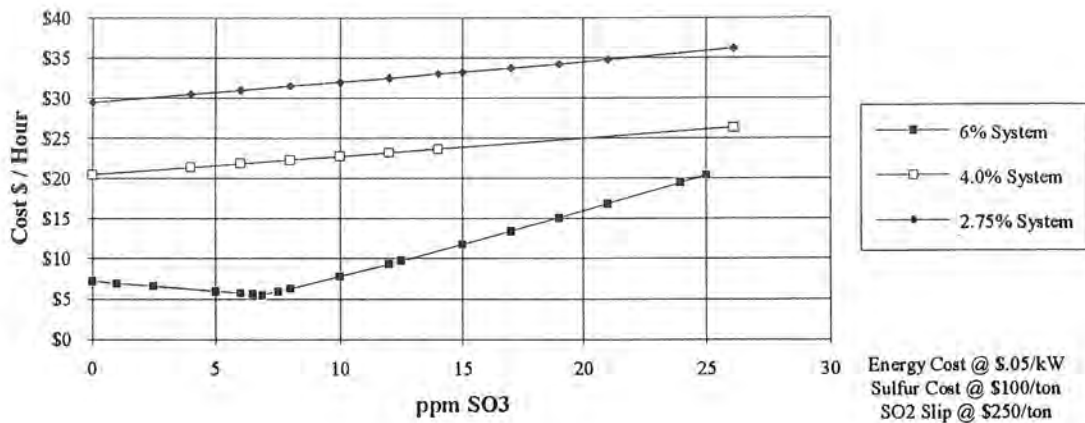


Figure 10 shows the projected operating cost comparison between the systems installed at Plant Bowen 3 and 4 and expected operating costs for systems operating at lower burner gas strengths. The two lower gas strength curves labeled "2.75% System" and "4.0% System" have constant sulfur burner air flows over the operating range.

**Figure 10**  
**Comparison of Gas Generator Operating Costs**  
**Bowen 3 & 4 vs. Predicted Gas Concentrations**



## Conclusions

1. The addition of the SO<sub>2</sub> cooler after the burner at plant Bowen lowered operating costs through reduced air flow rates. A 50% reduction in air flow into the burner was realized at maximum production rates. The cooler maintained the converter inlet temperatures within the desired range. It responded quickly to changes in sulfur and combustion air flow rates.
2. Gas conditioning systems that provide cooling for the burner outlet gases can achieve high converter efficiencies without being limited to 4% SO<sub>2</sub> at the burner outlet. These systems have lower operating costs, particularly at low production rates, than systems which maintain low sulfur burner gas strengths over the operating range.
3. Purchasing an SO<sub>3</sub> FGC system that is significantly larger than necessary for the expected operating rate increases the capital cost of the system and increases the operating cost. High concentration systems that can operate at high pressures reduce this risk because of their high turndown ratios.
4. Reducing the burner air flow with decreasing sulfur rates to maintain a constant ratio of burner air to sulfur in part allows the system to operate most

economically at about 50% of the design maximum capacity. This provides optimum operating costs at anticipated treat rates while providing sufficient margins if higher treat rates are needed.

5. The two stage converters are able to achieve more than 95% conversion, even with SO<sub>2</sub> gas concentrations greater than 6%.
6. Positive displacement blowers enable the system to operate with smaller vessels and less expensive hot gas pipe sizes. In addition the PD blowers have a significant reserve in pressure capability to maintain required air flow over time as the converter catalyst breaks down and the pressure drop increases.
7. FGC systems with low converter efficiency have increased usage of sulfur and increased SO<sub>2</sub> emissions up the stack. The difference between a system with 90% converter efficiency and 95% for a large boiler can be over \$25,000 per year in these costs.
8. High concentration SO<sub>3</sub> FGC systems may have operating costs as much as \$150,000 per year less than systems operating at low gas strengths.

## References

1. H.J. White. *Industrial Electrostatic Precipitation*. Addison Wesley Publishing Company, September 1962.
2. *BASF Catalyst*, BASF Catalog, issue date March 1984.
3. Robert Wright, David Woracek. *Advances in SO<sub>3</sub> Gas Plant Design and Control*. Presented at EPRI Ninth Particulate Control Symposium, October 1991.
4. William Archer, US Patent 3,993,429, *Gas Conditioning Means*, November 23, 1976.
5. Wilhelm Leusser, US Patent 4,931,265, *Conditioning Process for Dust-Containing Waste Gas*, June 5, 1990.