ABSTRACT

The goal of the study was to develop information and presentation materials demonstrating to current or potential new users of Illinois coal that a new generation of FGD system exists that is cost-effective, reliable, and able to control SO₂ emissions from power plants firing Illinois coal.

Nine types of FGD systems were surveyed. These nine systems were selected because they provide at least 95% SO₂ removal, have potential for lower cost than conventional scrubbing, and are either demonstrated or are part of a current demonstration program such as Clean Coal Technology or Combustion 2000. The technologies evaluated were: 1) Limestone wet scrubbing with forced oxidation, 2) Magnesium-enhanced lime wet scrubbing, 3) Dual-alkali wet scrubbing using limestone, 4) Advanced LIDS, 5) Gas suspension absorption, 6) SNRB process, 7) SNOX process, 8) NOXSO process, and 9) Copper-oxide dry scrubbing. Note that these systems include commercial and demonstration wet and dry scrubbing and combined processes including furnace injection.

The technologies were described; advantages and disadvantages discussed, and retrofit issues identified. Cost curves were developed for power plant sizes ranging from 100 MW to 500 MW. The costs for capital, fixed operation and maintenance (O&M), and variable O&M components were presented. The sensitivity of costs to sulfur contents for Illinois coals ranging from 2 to 5% were also shown. Other cost sensitivities evaluated included reagent costs, waste costs, and by-product credits. Vendors for these technologies and their experience were identified.

The cost data developed was applied to a Midwestern utility that is affected by Phase II of the Clean Air Act Amendments. The results showed that several FGD technologies could be considered, however, the utility can reduce the cost of FGD by 33% because they currently have a Limestone forced oxidation system at the station. The existing FGD system has spare capacity that can be utilized to reduce the cost of the FGD retrofit.
EXECUTIVE SUMMARY

Sales of Illinois coal in the year 2000, when the Clean Air Act Amendment of 1990 (CAAA) is in full effect, are projected to be at least 40% less than its sales in the mid-1990's. This loss of sales will result in the loss of many mining related jobs and premature closure of several mines. In order to improve the outlook for Illinois' coal mining industry, ICCI needs to be proactive in searching for technology innovation that will allow utilities, or others, to fire Illinois basin coal in new or existing power plants in an economic and environmentally safe manner.

The most significant negative characteristic of Illinois basin coal is its sulfur content ranging between 2.0 and 5.0% that leads to SO\textsubscript{2} generation ranging between 4 and 10 lb/MBtu. The effect of Clean Air Regulations is that a unit firing Illinois basin coal will need to remove at least 75% to 90% of the generated SO\textsubscript{2} and, in some situations, 95% removal might be necessary based on Best Available Control Technology (BACT) requirements. So to fire Illinois coal, the SO\textsubscript{2} needs to be controlled.

This project evaluated nine types of flue gas desulfurization (FGD) systems for control of SO\textsubscript{2} emissions. These nine systems were selected because they will be cost competitive with conventional limestone wet scrubbing and because they have either been commercially demonstrated or are being developed in current DOE programs. The technologies being evaluated are: 1) Limestone wet scrubbing with forced oxidation, 2) Magnesium-enhanced lime wet scrubbing, 3) Dual-alkali wet scrubbing using limestone, 4) Advanced LIDS, 5) Gas suspension absorption, 6) SNRB process, 7) SNOX process, 8) NOXSO process, and 9) Copper-oxide dry scrubbing.

This study described each FGD technology; quantified chemicals, utilities, and waste disposal required by each process; and developed an economic assessment of installing and operating each process at a 500 MW power plant firing Illinois coal. Cost sensitivities to plant size, coal sulfur variation, and variation in other cost elements were also determined. Flow schematics illustrating the processes and graphs illustrating the economic assessments were prepared.

The results of the survey were a final report and presentation materials which can be used by ICCI to inform customers that use of Illinois basin coals while meeting SO\textsubscript{2} emissions goals is economical and environmentally sound, and can be competitive with other approaches such as coal switching.
OBJECTIVE

The overall goal of this project was to demonstrate to current or potential new customers of Illinois coals that a new generation of FGD systems exists that is cost-effective, reliable, and able to control sulfur emissions from the firing of Illinois basin coals. This was done by surveying nine (9) flue gas desulfurization (FGD) system types and comparing their technical attributes and costs when applied to steam electric generating units firing a typical high-sulfur, high-chloride Illinois basin coal.

Specific objectives of this FGD survey were to:

1. Prepare descriptions of the current state of each FGD technology discussing how the process functions and the process chemistry involved, identifying components included with the system, examining retrofit issues, listing advantages and disadvantages, and presenting the commercial status and vendors offering the process;

2. Quantify the chemical and utilities required by each process and the wastes or byproducts generated by each process for a typical 500 MW, 3% sulfur coal-utility power plant application;

3. Estimate the capital costs for new installations of each technology at a typical 500 MW, 3% sulfur coal utility power plant application, the fixed and variable costs of operating the systems, and the levelized costs showing the economic impacts on the cost of electricity; and

4. Evaluate the sensitivities of process economics to variations in unit sizes ranging from 100 MW to 500 MW, variations in coal sulfurs ranging from 2% to 5%, and variations in reagent prices, waste disposal costs, byproduct values, or other cost elements.

The results of the survey were a final report and presentation materials which can be used by ICCI to inform customers that use of Illinois basin coals while meeting SO₂ emissions goals is economical and environmentally sound, and can be competitive with other approaches such as coal switching.

INTRODUCTION AND BACKGROUND

Sales of Illinois coal in the year 2000, when the Clean Air Act Amendment of 1990 (CAAA) is in full effect, are projected to be at least 40% less than its sales in the mid-1990's. This loss of sales will result in the loss of many mining related jobs and premature closure of several mines. In order to improve the outlook for Illinois' coal mining industry, ICCI needs to be proactive in searching for technology innovation that
will allow utilities, or others, to fire Illinois basin coal in new or existing power plants in an economic and environmentally safe manner.

The most significant negative characteristic of Illinois basin coal is its sulfur content ranging between 2.0 and 5.0% that leads to SO₂ generation ranging between 4 and 10 lb/MBtu. The effect of Clean Air Regulations is that a unit firing Illinois basin coal will need to remove at least 75% of the generated SO₂ and, in many situations, 90% to 95% removal might be necessary. So to fire Illinois coal, the SO₂ needs to be controlled.

This project evaluates flue gas desulfurization methods for controlling SO₂ emissions. FGD technology has advanced since the first systems were installed in the 1970's with current processes offering improved reliability and reduced costs. Advanced FGD systems are being used to meet Phase I Clean Air Act Amendments, and more are being demonstrated through Department of Energy (DOE) programs. Nine (9) of these FGD processes capable of 95% SO₂ removals were selected for evaluation and comparison.

The technologies chosen included processes which control only SO₂ by wet scrubbing methods (Limestone Forced Oxidation, Magnesium-enhanced Lime, and Limestone Dual Alkali) or by dry scrubbing methods (Limestone Injection with Dry Scrubbing and Gas Suspension Absorption), and process which control both SO₂ and NOₓ by dry scrubbing methods (SNRB) or by catalytic/regenerable sorbent methods (SNOX, NOXSO, and Copper Oxide).

These nine FGD types were selected because they will be cost competitive with conventional limestone FGD and they have either been commercially proven and available or are currently being developed in the DOE Clean Coal Technology program or the DOE Combustion 2000 program for near-term commercialization.

PROCEDURES

This project is an engineering survey and economic assessment of nine current FGD technologies which would enable the use of high-sulfur Illinois coal while complying with the Clean Air Act (CAA). The study relied on existing industry data; further laboratory or full-scale testing was beyond the scope of the study. A design basis was developed for applying the technologies to a 500 MW net output pulverized coal power plant firing Illinois #6 coal (3% sulfur and 0.2% chloride). Design desulfurization performances were 95% removal of SO₂.

Economic evaluations were done for these applications by estimating the capital costs, the fixed operating and maintenance costs, the variable operating costs, and the operating credits from byproduct sales or heat recovery. All estimates were based on a 65% capacity factor. These costs were also adjusted to determine costs for 100 MW and 300 MW size
units and to coal sulfurs of 2, 4, and 5%. Levelized costs were also prepared from this data using a constant dollar basis.

A case study was also prepared with this information by applying the technologies to an existing midwestern utility in an evaluation of FGD options.

RESULTS AND DISCUSSION

The following summaries describe the nine FGD technologies investigated in this study and present technical and commercial assessments. More detailed technical writeups for each technology are available individually from ICCI as separate documents.

FGD Technology Descriptions

Limestone Forced Oxidation (LSFO)

Limestone Forced Oxidation is a wet scrubbing process and is the flue gas desulfurization technology most frequently selected in the 1990's for utility boilers. The FGD system, shown schematically in Figure 1, is located between the baghouse or electrostatic precipitator and the stack. LSFO uses a limestone reagent to produce gypsum solids either for disposal or for sale. Further processing is required to produce commercial-grade gypsum for sale.

Flue gas is treated in an absorber by contacting the gas with a limestone slurry. The slurry absorbs $\text{SO}_2$ from the gas and reacts the $\text{SO}_2$ with limestone in the slurry to form calcium sulfite. The gas is also cooled and saturated with water from evaporation of some of the slurry. The desulfurized flue gas then passes through mist eliminators to remove entrained droplets of slurry and is sent to the stack.

After contacting the gas, the slurry collects in a reaction tank at the bottom of the absorber and is recycled back to the absorber. Slurry in the reaction tank is aerated to oxidize calcium sulfite and convert it to gypsum solids. This is where the term "forced oxidation" originates and it distinguishes this process from older, more troublesome "natural oxidation" technology. Aeration forces the precipitation to occur in the reaction tank on existing crystals and prevents uncontrolled oxidation in the absorber from forming a hard, adherent scale on surfaces of absorber internals as happened with early generation wet FGD systems. Fresh limestone slurry is also added to the reaction tank to replenish limestone consumed in the process.

A portion of the slurry is withdrawn to remove the collected gypsum. The solids are concentrated in a thickener or hydroclone and then further dewatered with a centrifuge or a vacuum filter. This produces a cake of gypsum solids which is landfilled either alone
or blended with flyash. If there is a nearby market for gypsum, additional equipment can be added to produce a purified gypsum for sale instead.

Continuing development of the LSFO process includes improvements such as grinding the limestone reagent to a finer size, reducing the circulation rate of slurry (liquid to gas ratio, known as L/G), and use of smaller, higher-velocity absorbers. These advancements can reduce capital costs by 15%, reduce operating costs, and reduce consumption of electrical power.

LSFO systems have achieved SO₂ removal efficiencies as high as 99% in a variety of power plants firing a variety of fuels. The higher removals are typically attained through the use of additives such as dibasic acid in the recirculating slurry or by recirculating larger quantities of slurry through the flue gas.

Limestone Forced Oxidation FGD systems are in operation at many facilities ranging in size from less than 100 MW to 750 MW and are under construction at many more. Applications include commercial units with coal sulfurs as high as 4.5%. Systems are commercially available from a number of vendors including ABB Environmental Systems, Babcock & Wilcox, General Electric Environmental Systems; Noell, Wheelabrator Air Pollution Control, Pure Air, and Saarburg-Holter.

**Magnesium Enhanced Lime Scrubbing (MgLime)**

Magnesium Enhanced Lime Scrubbing is a wet scrubbing process derived from hydrated lime scrubbing technology. The process was developed to improve SO₂ removal efficiencies and minimize scaling tendencies in early-generation lime scrubbing technology. The FGD system, shown schematically in Figure 2, is located between the baghouse or electrostatic precipitator and the stack. MgLime uses a magnesium-enhanced lime reagent to produce waste calcium sulfite solids for disposal.

Flue gas is treated in an absorber by contacting the gas with a slurry of slaked quicklime. The slurry absorbs SO₂ from the gas and reacts the SO₂ with hydrated lime in the slurry to form calcium sulfite. The gas is also cooled and saturated with water from evaporation of some of the slurry. The desulfurized flue gas passes through mist eliminators to remove entrained droplets of slurry and is then sent to the stack.

The process is similar to a conventional lime scrubber, but uses a lime enhanced with magnesium. The magnesium content allows the slurry to have both a higher pH and a lower calcium content than for conventional lime slurries. This reduces the rate of calcium sulfite oxidation in the absorber and minimizes the formation of gypsum scale which might otherwise occur as a result of natural oxidation. Forced oxidation is not needed in magnesium enhanced lime scrubbing systems for control of gypsum scaling. It also increases the rate of SO₂ absorption, permitting smaller equipment sizes compared to limestone systems.
After contacting the gas, the slurry collects in a reaction tank the bottom of the absorber and is recycled back to the absorber. A portion of the slurry is withdrawn to remove the collected calcium sulfite solids. This slurry is dewatered in a thickener and vacuum filter system to produce a cake which must then be stabilized with flyash or lime for disposal of the waste calcium sulfite in a landfill.

Lime consumed in the process is replenished by adding fresh lime slurry to the reaction tank. Magnesium remains dissolved in the slurry and is not chemically consumed by the process, but losses due to liquid leaving as moisture with the waste calcium sulfite cake must be replaced. Quicklime with the necessary magnesium content, known as thiosorbic lime, is available from Dravo or high-calcium quicklime supplemented with dolomitic lime can be used.

MgLime systems have achieved SO$_2$ removal efficiencies as high as 98% in a variety of power plants firing a variety of fuels. The higher removals are typically attained by recirculating larger quantities of slurry through the flue gas.

Magnesium Enhanced Lime FGD systems are in operation at many facilities ranging in size from 41 to 1400 MW. Applications include commercial units with coal硫urs as high as 5.1%. Most of these were installed in the 1970's and early 1980's, but several have been installed in the 1990's or are now being constructed. Systems are available from many U.S. vendors including ABB Environmental Systems, Babcock & Wilcox, General Electric Environmental Systems, Noell-KRC, Wheelabrator Air Pollution Control, Pure Air, Bechtel National, Joy, Dravo, and Riley-Deutsche Babcock.

**Limestone Dual Alkali (LSDA)**

Limestone Dual Alkali is a wet scrubbing process developed to prevent gypsum scaling of absorber internals while still retaining the operating cost advantages of an inexpensive limestone reagent. The FGD system, shown schematically in Figure 3, is located between the baghouse or electrostatic precipitator and the stack. LSDA uses a limestone reagent to produce waste calcium sulfite solids for disposal.

Flue gas is treated in an absorber by contacting the gas with a recirculating solution of sodium sulfite. The solution absorbs SO$_2$ from the gas and reacts the SO$_2$ with sodium sulfite to form a soluble reaction product in the solution. The gas is also cooled and saturated with water from evaporation of the solution. Desulfurized flue gas passes through mist eliminators to remove entrained droplets of solution and is then sent to the stack.

Because the scrubbing reagent is a solution rather than a slurry as in limestone scrubbing, the potential for fouling absorber internals with solids is minimal. Further, unlike limestone scrubbing, any natural oxidation of sulfite in the absorber forms a soluble sodium sulfate product instead of gypsum scale. Since the sodium sulfite solution does
not require time to dissolve, it is more reactive than limestone slurry. The increased rate of SO$_2$ absorption permits smaller equipment sizes than for limestone scrubbing.

A portion of the recirculating solution is continuously withdrawn for regeneration with limestone in a separate regeneration system. In this system, the sodium sulfite solution is reacted with limestone slurry in a mixed tank to precipitate calcium sulfite solids and to recover the sodium sulfite for reuse. The resulting slurry of waste calcium sulfite solids and regenerated sodium sulfite solution is dewatered in a thickener and vacuum filter system, and the clear recovered sodium sulfite solution is returned to the absorber. The cake of waste calcium sulfite solids is then stabilized with flyash or lime for disposal in a landfill.

The net result of the process is consumption of limestone and removal of SO$_2$ as waste calcium sulfite solids. This is much like the overall result of limestone scrubbing. Sodium sulfite is not consumed in the process, but losses due to liquid leaving as moisture in the waste calcium sulfite must be replaced by adding soda ash solution to the absorber. The use of two reagents, a sodium-based alkali for contacting the gas and a calcium-based alkali for removing absorbed SO$_2$ distinguishes the LSDA process from single-alkali processes.

The performance history of Dual Alkali systems utilizing limestone is limited to full-scale piloting and to oil-fired boilers overseas, but similar systems based on lime rather than limestone have achieved SO$_2$ removals of 90 to 98% in a variety of power plants. Performances with limestone should be similar to those achieved with lime.

There are no commercial dual alkali FGD systems utilizing limestone operating in the U.S., but similar systems based on lime rather than limestone are in operation at five U.S. facilities ranging in size from 265 to 575 MW. In Japan, there are ten commercial installations of limestone-based systems using similar Japanese technology at facilities up to 450 MW. These commercial applications include units with coal sulfurs as high as 4%. There is no U.S. supplier of limestone dual alkali systems, but process design is available from Codan Associates and Central Illinois Public Service. Ontario Hydro now owns the patent and U.S. license for dual alkali FGD technology.

**Limestone Injection with Dry Scrubbing (LIDS)**

Limestone injection with dry scrubbing is a dry desulfurization process combining dry scrubbing (spray dryer) technology with limestone injection at the furnace. The dry scrubber reactor where desulfurization takes place is located between the air heater and the baghouse or electrostatic precipitator. The LIDS process, shown schematically in Figure 4, uses limestone to produce dry waste calcium sulfite solids for disposal.

Flue gas is desulfurized in a cocurrent reactor by spraying a concentrated slurry of lime and flyash solids into the flue gas stream which then passes down through the reactor.
Water in the slurry evaporates in the reactor, cooling and humidifying the gas. The slurry spray absorbs SO\(_2\) from the flue gas, reacts the SO\(_2\) with lime in the slurry, and simultaneously dries as it passes down through the reactor. This converts the absorbed SO\(_2\) into dry calcium sulfite particles suspended in the gas. At the bottom of the reactor, coarser lime and calcium sulfite particulates drop out while finer particulates remain in the gas.

The gas exiting the dry scrubber reactor next flows to a baghouse or electrostatic precipitator where the remaining calcium sulfite and unreacted lime particles are removed along with flyash in a dry form. Residual SO\(_2\) in the gas continues to react with unreacted lime as the gas passes through the collected solids, although at a slower rate than in the dry scrubber reactor. Not all the remaining lime is reacted, so the flyash solids contain unreacted lime as well as the calcium sulfite solids generated in the FGD process.

The coarse waste solids collected at the bottom of the dry scrubber reactor plus a portion of the dry flyash and FGD solids mixture collected in the baghouse or electrostatic precipitator is recovered and mixed with water. The resulting slurry, containing a mixture of hydrated lime and flyash, is sprayed into the dry scrubber reactor. This recycles some of the unreacted lime in the solids and improves reagent utilization rates.

The remainder of the solids from the baghouse or electrostatic precipitator is landfilled as a dry waste. Because calcium sulfite and unreacted lime solids are mixed with flyash, there is no potential for recovery of waste reaction products as a saleable byproduct. The mixture of FGD solids with the flyash also prevents the commercial sale of flyash.

Lime consumed by the reaction with SO\(_2\) in the dry scrubbing process is not added at the dry scrubber, but is provided by pneumatically injecting pulverized limestone into the furnace. The primary purpose of injection at this location is to calcine the limestone to quicklime, a more reactive sorbent. This enables the use of a lower cost limestone reagent without sacrificing reactivity. Limestone consumption is higher than for wet limestone scrubbing, but the greater reagent cost is offset by simpler system operation.

Operating experience for the combined limestone injection with dry scrubbing process is limited, but experience with similar dry scrubbing technology indicates SO\(_2\) removals of 70 to 95% can be achieved with high sulfur coals at a 20 to 40\(^\circ\) F adiabatic approach to saturation. Closer approaches to saturation result in greater SO\(_2\) removal efficiencies, but at the risk of causing accumulations of damp solids in the dry scrubber reactor.

There are no full-scale commercial installations of the LIDS combination of spray dryer and furnace sorbent injection technology, although the process is commercially available from the developer Babcock & Wilcox. Individual application of spray dryer technology is well proven for coal sulfurs up to 2% and applications of spray dry technology to higher sulfur coals has been tested in pilot and full scale systems. There are also
installations using high sulfur coal in Europe. Furnace limestone injection has been demonstrated in test programs using a variety of coals in the U.S., Canada, Europe, and Japan.

**Gas Suspension Absorption (GSA)**

Gas Suspension Absorption is a dry scrubbing process developed to provide high desulfurization efficiencies and low reagent consumptions not possible with conventional spray dryers. The dry scrubber reactor where desulfurization takes place is located between the air heater and the baghouse or electrostatic precipitator. The GSA process, shown schematically in Figure 5, uses pebble quicklime to produce dry calcium sulfite solids for disposal.

Flue gas is desulfurized in a cocurrent vertical reactor by spraying a slurry of slaked quicklime into the flue gas stream at the bottom of the reactor. The flue gas then carries the injected slurry up through the reactor. A dual-fluid injection nozzle combining air and lime slurry in the nozzle is used to atomize the slurry into fine droplets which evaporate quickly. The evaporating water cools and humidifies the gas and dries the injected solids to less than 1% moisture.

In the reactor, the slurry absorbs $\text{SO}_2$ as it dries, and the $\text{SO}_2$ reacts with the lime to form dry calcium sulfite particles suspended in the flue gas. Removals are enhanced by operating near the gas saturation temperature, so water is added with the lime slurry at a controlled rate to maintain the gas within about 10 to 20$^\circ$ F of saturation temperature. Because the solid particles are dried to such low moisture contents, solids do not build up on reactor walls even at these close approaches to saturation.

Desulfurized gas and suspended particulates exiting the top of the reactor then flow to a cyclone separator where calcium sulfite and unreacted lime particles are removed in a dry form. Most of the FGD solids in the gas stream, about 99%, are recycled to the reactor inlet just above the slurry injection nozzle using a screw conveyor. This reuses unreacted lime present in the collected solids and maintains a high solids concentration within the reactor. The resulting bed of solids suspended in the reactor aids in $\text{SO}_2$ capture, solids drying, and prevention of scaling on internal surfaces by providing intimate, turbulent contact with the flue gas.

Fine calcium sulfite and unreacted lime particulates escaping the cyclone separator are captured along with flyash in the downstream electrostatic precipitator or baghouse. Residual $\text{SO}_2$ in the flue gas continues to react with the alkaline solids in either particulate collection device, although removals are 3 to 5 percentage points higher in applications using a baghouse.

The mixture of flyash, calcium sulfite, and excess lime collected at the particulate control device and excess solids collected at the cyclone are combined for disposal as a dry
waste. Because calcium sulfite and unreacted lime solids are mixed with flyash, there is no potential for recovery of waste reaction products as a saleable byproduct. The mixture of FGD solids with the flyash also prevents the commercial sale of flyash.

Operating experience for GSA systems applied to boiler applications is limited to pilot testing with 2.6 to 3.5% sulfur western Kentucky coals. Test results demonstrated SO$_2$ removals up to 99% and indicated SO$_2$ removals over 90% can consistently be achieved at moderate lime consumptions. Decreasing the approach to saturation temperature (tested as low as 8° F) and increasing coal chlorides (tested up to 0.12%) resulted in higher removals.

The GSA process was developed by FLS Industries in Denmark and has been commercially applied to nine Danish municipal waste incinerators to capture SO$_2$ and HCl emissions. In the U.S., a 10 MW pilot program was recently completed at a 150 MW commercial power plant with test durations totaling 7700 hours of operation. The first commercial U.S. application has been ordered for a 50 MW boiler to enable firing high-sulfur Ohio coal. The process is commercially available in the U.S from AirPol.

**SNRB Process**

The SNRB (SO$_x$/NO$_x$/RO$_x$/Box) process combines removal of SO$_2$ and reduction of NO$_x$ in a high-temperature baghouse located between the economizer and the air heater. It is a dry desulfurization process which uses hydrated lime to produce waste calcium sulfite solids for disposal. Ammonia is also used for selective catalytic reduction (SCR) of NO$_x$ inside the baghouse. The SNRB process is shown schematically in Figure 6.

Dry hydrated lime sorbent is pneumatically injected into the flue gas in the ductwork between the economizer and a high-temperature baghouse. The optimum injection point has been found to be directly ahead of the baghouse rather than in higher temperature regions further upstream. The mixture of sorbent and flyash particulates is then removed in a pulse-jet baghouse using ceramic fiber bags and operating at a temperature of 800 to 850° F. In the baghouse, the collected particulates form a layer of alkaline solids on the filter bags.

While some of the lime reacts with SO$_2$ to form solid calcium sulfite particles in the ductwork, most of the desulfurization reaction takes place inside the baghouse as the flue gas passes through the layer of solids which has been collected on the filter bags. The mixture of excess lime sorbent, calcium sulfite, and flyash captured in the baghouse is landfilled for disposal. There is no potential for recovery of the waste solids as a saleable byproduct, since the FGD solids are mixed with flyash. The mixture of calcium sulfite and unreacted lime with flyash also prevents commercial sale of flyash.

Because SO$_2$ is removed in a high-temperature baghouse ahead of the air heater, the potential for condensation of sulfuric acid when the flue gas is cooled is lessened. This
permits a lower flue gas outlet temperature in the air heater and can improve thermal efficiencies. The process developer projects that outlet temperatures can be lowered by 20 to 100° F, for a 1% to 3% improvement in power system efficiency.

NO\textsubscript{x} is also removed inside the baghouse by using selective catalytic reduction (SCR). Ammonia vapor is injected in the ductwork ahead of the baghouse to reduce NO\textsubscript{x} to nitrogen and water vapor over SCR catalyst installed within the filter bags. This catalyst location protects the catalyst from particulate erosion or fouling and the high baghouse operating temperature allows catalytic reduction without requiring flue gas reheating to reaction temperatures.

Lime sorbent used in the process is prepared by hydrating pebble quicklime in an atmospheric hydrator. Just enough moisture is added to completely hydrate the available calcium oxide content of the quicklime while still producing the hydrate as a dry powder. Alternatively, commercial hydrated lime can be purchased; but it is more costly and more expensive to ship than pebble lime.

The process has been demonstrated in pilot testing with 3.4% sulfur coal to be capable of removing over 80% of flue gas SO\textsubscript{2} using commercial hydrated lime. Higher removals, 85 to 90%, were attained by increasing hydrated lime feed rates. Use of modified hydrated limes (sugar hydrated lime and lignosulfonate hydrated lime) also achieved over 90% removals. Extending these results suggests 95% SO\textsubscript{2} removal performances can be achieved with further increases in lime usage.

Particulate emissions were consistently below the NSPS standard of 0.03 pound per million Btu and averaged 0.018 pound per million Btu. This corresponded to a collection efficiency of 99.89%.

The process was developed as a proprietary process by Babcock & Wilcox and is in the pilot stage of development. A 5 MW pilot program utilizing a flue gas slipstream from a 150 MW commercial plant was recently completed. Test durations totaled 2300 hours of operation. Commercialization is proceeding based on the successful pilot testing with initial efforts being focused toward smaller units where cost advantages may be greatest.

**SNOX Process**

The SNOX process is a dry catalytic process for removing both NO\textsubscript{x} and SO\textsubscript{2} from flue gas. The system, shown schematically in Figure 7, is located downstream of the air heater. It utilizes ammonia injection with selective catalytic reduction (SCR) to convert NO\textsubscript{x} to nitrogen gas and a separate oxidation catalyst and condenser to recover SO\textsubscript{2} as a saleable sulfuric acid byproduct. Reagents are not required for flue gas desulfurization.

Flue gas from the air heater is first treated in a high-efficiency baghouse to remove particulates which could foul or poison the SNOX desulfurization catalyst in the
downstream SO₂ converter. A reverse-air or pulse-jet baghouse equipped with Goretex teflon membrane bags is used to achieve the required cleanliness. This protects the SCR catalyst as well, minimizing any attrition or pluggage and extending catalyst life.

The flyash-free gas is then reheated to process reaction temperature with a gas-to-gas heat exchanger using heat from treated flue gas leaving the process reactors. The hot gas first flows through a selective catalytic reduction (SCR) reactor where NOₓ is converted to nitrogen gas and water vapor. A high activity, titanium oxide-based catalyst is used and operating temperatures are 650 to 800° F. Ammonia needed for the reaction is injected into the duct ahead of the reactor.

The gas is then sent to an oxidation reactor. Because the reaction temperature for efficient catalytic oxidation of SO₂ is 790 to 810° F, a support burner firing natural gas is used to raise the temperature of the SCR reactor outlet gas. In the oxidation reactor, most of the SO₂ is converted to SO₃ using a sulfuric acid catalyst. Ammonia slip from the SCR reactor is oxidized, and a portion of NOₓ remaining in the flue gas or formed from ammonia slip is also oxidized to NO₂ by the catalyst. As a result, NOₓ removal efficiencies ahead of the oxidation reactor must be maintained above 90% to prevent excessive NO₂ levels and formation of a brown plume at the stack.

Hot gas leaving the SO₂ oxidation reactor is used to heat the incoming flue gas in the gas-to-gas heat exchanger. The partially cooled gas, still above its acid dew point, then flows to a falling film condenser (called the WSA tower) to recover sulfuric acid. There, the gas is air-cooled to about 210° F and sulfuric acid condenses as a falling film inside glass tubes. This produces a concentrated acid at a commercial strength of 93 to 98% H₂SO₄ which is recovered for sale as a by-product.

The desulfurized flue gas passes through a demister to remove entrained droplets of sulfuric acid mist and is sent to the stack at a 210° F temperature. The hot air exiting the WSA condenser is at about 400° F and is sent to the boiler for heat recovery.

All the SO₂ removed from the flue gas is converted to commercial-grade sulfuric acid for sale, so there is no waste stream generated by the process. However, particulates which pass through the baghouse accumulate on the sulfur oxidation catalyst and must be periodically removed by screening the catalyst. The resulting waste dust consists primarily of flyash, but also includes fines from attrition of the catalyst. Occasional replacement of SCR and oxidation catalysts is also required, but the spent catalysts can be returned to the manufacturer for recovery of metals.

The process has achieved SO₂ removal efficiencies of 90 to 95% in full scale operation in a commercial installation in Denmark and 96% in pilot testing in the U.S. Coal sulfurs were 0.5 to 3.0 % in Denmark and 3.4% in the U.S. pilot test.
The process was developed by Haldor Topsoe and is commercially available through ABB Environmental Systems in the U.S. There are currently two commercial installations; a 305 MW coal-fired plant in Denmark, and a 30 MW plant in Italy firing petroleum coke. Both systems have been in operation since 1991.

In the U.S., a pilot plant has been installed on a 35 MW equivalent flue gas slipstream from a 108 MW unit. This system has been operated for over 7,800 hours and has produced over 5,400 tons of commercial-grade sulfuric acid. The acid has been sold to the fertilizer industry and to the steel industry, and has also been used internally for water treatment.

**NOXSO Process**

The NOXSO process is a dry adsorption process which removes both NO\textsubscript{x} and SO\textsubscript{2} from flue gas using a regenerable sorbent. The system, shown schematically in Figure 8, is located between the air heater and the baghouse or electrostatic precipitator. Reagents are not required for removal of either SO\textsubscript{2} or NO\textsubscript{x}, since the process is based on adsorption and desorption. NO\textsubscript{x} is recycled to the furnace for destruction, while SO\textsubscript{2} is recovered as a saleable sulfuric acid byproduct. Alternatively, saleable elemental sulfur can be produced as a byproduct instead of acid.

Flue gas leaving the air heater is treated in a multi-stage fluidized bed adsorber. In the adsorber, spherical beads of sorbent contact the gas and adsorb both NO\textsubscript{x} and SO\textsubscript{2} from the gas. Sorbent flows continuously through the adsorber and spent sorbent is sent to a separate regeneration system. The flue gas is also cooled within the adsorber using water sprays. The evaporating water spray cools the gas and maintains the adsorber operating temperature at 250° F. All of the water spray evaporates in the adsorber, so there is no liquid discharge stream. Treated gas flows through a cyclone for removal of attrited sorbent particles and is then sent to the plant's baghouse or electrostatic precipitator.

Spent sorbent containing adsorbed NO\textsubscript{x} and SO\textsubscript{2} is recovered in a regeneration system. Attrited sorbent removed at the cyclone is also added to the spent sorbent for processing in the regeneration system. There, the sorbent is first heated in a sorbent heater to desorb NO\textsubscript{x}. Hot air, heated to 1325° F with a natural gas-fired air heater, is used to strip NO\textsubscript{x} from the sorbent. The heat from the air stream raises the sorbent temperature to 1150° F. At this temperature, all the adsorbed NO\textsubscript{x} desorbs into the air stream.

The air stream containing the desorbed NO\textsubscript{x} leaves the sorbent heater at 445° F, is cooled to 150° F by heating a slipstream of boiler feedwater, and is recycled to the boiler by mixing it with combustion air. This recycling destroys most of the NO\textsubscript{x} in the furnace and suppresses formation of additional NO\textsubscript{x} to equilibrium levels. It also recovers the heat used in the regeneration process. The recycle NO\textsubscript{x} stream contains attrited sorbent which is also returned to the boiler and ultimately collected with bottom ash or flyash.
The hot sorbent from the sorbent heater is next treated in a countercurrent moving-bed regenerator to desorb SO\textsubscript{2}. Natural gas is fed to the upper section of the regenerator to react with the SO\textsubscript{2} at elevated temperatures. After the sorbent leaves this zone it is then contacted with low pressure steam in the lower section of the regenerator to remove the remaining SO\textsubscript{2}. The sorbent regeneration process produces an offgas with high concentrations of SO\textsubscript{2} and H\textsubscript{2}S. Some elemental sulfur vapor is formed as well.

Finally, the regenerated sorbent is cooled in a four-stage, fluidized bed sorbent cooler. Ambient air is used to reduce the sorbent temperature to 275\textdegree F. This heats the air to 850\textdegree F, and the heat from this step is recovered by reusing the air for sorbent heating in the first regeneration step. The cooled sorbent is returned to the adsorber, completing the cycle.

Offgas from the regenerator containing desorbed SO\textsubscript{2} is processed in a sulfur recovery unit for conversion to saleable byproduct. A contact sulfuric acid plant can be used to produce sulfuric acid, or a Claus plant can be used to produce elemental sulfur. All the SO\textsubscript{2} removed from the flue gas is recovered, so there is no waste stream of sulfur reaction products.

Bench-scale and pilot testing of the NOXSO process indicates that 96\% SO\textsubscript{2} removals are achievable with the process. NO\textsubscript{x} removals of 87\% were also demonstrated in these tests. Testing has been performed with 3 to 5\% sulfur Ohio coal (3.5 \% average), and has been performed with SO\textsubscript{2} spiking of flue gas to simulate coal sulfurs as high as 6\%.

The NOXSO process is being developed by NOXSO Corporation and MK-Ferguson Company and is entering the full-scale stage of commercialization. A 5 MW pilot demonstration has been completed, and two commercial demonstration projects, one in the U.S. and one in Denmark, are now underway. The U.S. system is a 150 MW coal-fired boiler application designed for 98\% SO\textsubscript{2} removal, while the Denmark project is a coal-fired heat/power station generating 12 MW heat and 5 MW electricity.

**Copper Oxide Process**

The Copper Oxide process is a dry adsorption process which removes both SO\textsubscript{2} and NO\textsubscript{x} from flue gas using a regenerable sorbent. The system, shown schematically in Figure 9, is located between the economizer and the air heater. Reagent is not required for SO\textsubscript{2} removal, since the process is based on adsorption and desorption. SO\textsubscript{2} is recovered as a saleable sulfuric acid byproduct or, alternatively, as a saleable elemental sulfur byproduct. Ammonia injection is used to catalytically reduce NO\textsubscript{x}.

Flue gas from the economizer is desulfurized by passing it through a moving bed of sorbent beads in a cross-flow adsorber. Spherical beads of an alumina substrate impregnated with copper oxide move downward as narrow beds contained within vertical screens or louvers. Sorbent flow is continuous, with fresh sorbent added at the top of the
adsorber and spent sorbent removed at the bottom. Flue gas passes horizontally through the moving beds of sorbent and the treated gas exiting the adsorber is sent to the air heater.

The primary function of the sorbent is to remove SO₂ from the hot flue gas. This occurs most efficiently at higher temperatures, so the flue gas is treated at the economizer exit temperature of 750° F. The adsorption reaction raises the gas temperature an additional 70 to 80° F. At these temperatures, carbon steel can still be used for adsorber internals.

The sorbent beads are also an effective catalyst for removing NOₓ. By injecting ammonia upstream of the adsorber, NOₓ is reduced to nitrogen and water vapor by selective catalytic reduction (SCR) reactions. Simultaneous NOₓ reduction is not necessary for the desulfurization process, however, and the adsorber can be operated for SO₂ adsorption alone if NOₓ reduction is not needed.

Additionally, the moving bed acts as a filter and removes much of the flyash from the inlet flue gas stream. Removals are in the 50 to 90% range, so further particulate removal is still required in a downstream baghouse or electrostatic precipitator. Flyash removals in the adsorber reduce ash loadings in the air heater and the particulate removal device, however, improving performances of these systems.

Spent sorbent removed from the bottom of the adsorber for regeneration is first screened to separate flyash from the sorbent beads. Flyash from this step is conveyed to the flyash silo where it is combined with other flyash captured by the particulate removal equipment. The cleaned sorbent is then conveyed to a separate vessel for regeneration.

In the regeneration system, the spent sorbent is contacted with a countercurrent flow of a reducing gas to desorb SO₂ which was removed from the flue gas. The sorbent flows downward as a moving bed in the regeneration vessel and the reducing gas flows upward through the descending bed. This recovers the SO₂ in a concentrated offgas stream. Regenerated sorbent is returned to the moving bed adsorber.

The preferred reducing gas, a mixture of hydrogen and carbon monoxide, is produced from natural gas by heating a mixture of natural gas and steam to 1600° F in a natural gas fired reformer. The hot reducing gas produced by the reforming reaction serves to heat the sorbent to the 800° F temperature needed for regeneration as well as to desorb SO₂ in the regeneration step.

Alternatively, natural gas can be used directly as the reducing gas. Direct regeneration with natural gas does not require a reformer, but does require a larger regenerator since methane is a less reactive reducing gas. It also requires preheating of the sorbent in a separate gas fired heater prior to regeneration.
Offgas from the regenerator is processed in a sulfur recovery unit for conversion of the recovered SO₂ to saleable byproduct. A contact sulfuric acid plant can be used to produce sulfuric acid, or a Claus plant can be used to produce elemental sulfur. All the SO₂ removed from the flue gas is recovered, so there is no waste stream of sulfur reaction products.

The Copper Oxide process chemistry has been well established with fixed or fluidized beds and the moving bed adsorber design has been demonstrated with other sorbents, but testing of the combination of copper oxide sorbent in a moving bed contactor is limited to bench scale tests. Results of these tests along with performances of analogous systems indicate SO₂ removals of up to 99% can be attained. Pilot scale testing with 2.6% sulfur Illinois coal to confirm the process began in April, 1995.

The process is in the early stage of development and is not commercially available as yet. The U.S. Department of Energy holds the patent for the moving-bed adsorber developed by Rockwell International and is developing the process through its Pittsburgh Energy Technology Center and various subcontractors. Individual process elements have been investigated in bench-scale testing, and a pilot-scale Life Cycle Test System with a 40 lb/hr pulverized coal combustor is now being used to further develop process parameters.

**Retrofit Issues**

FGD process absorbers and reactors are large and must be integrated into the flue gas path. Sufficient space is usually available only beyond the existing stack or to one side of an existing unit. The expense of ducting flue gas to such a location and the complexity of installing flue gas tie-ins are added costs for a retrofit project. Some of the processes also require additional tie-ins to the combustion air path or the furnace, further increasing installation complexity.

Additional space is also required for support facilities such as solution or sorbent regeneration, chemical handling, waste dewatering, or byproduct production. These facilities can require much more space than the desulfurization vessels, but can be located away from the main unit where there is less congestion.

Wet scrubbing processes saturate the desulfurized gas stream with moisture which can lead to corrosive conditions. This is especially true with high chloride coals such as those sometimes found in Illinois and elsewhere. As a result, FGD equipment must be designed for corrosive service. Downstream ductwork is also subjected to corrosive conditions and must be protected as well. The chimney will operate as a wet stack and is usually replaced with a new, acid-resistant chimney, but an existing steel liner could be wallpapered with an alloy and reused.

Dry scrubbing processes operate near the flue gas dew point, so there is a potential for condensation and corrosion on cool surfaces in downstream ductwork and equipment.
Additional thermal insulation may be required to prevent localized condensation in particulate collection devices. Protection of ductwork with a corrosion-resistant coating may also be required.

FGD systems impose an additional pressure drop in the flue gas path, which affects ID fan performance. Most of the technologies increase pressure drops by 4 to 10 inches, but some have still larger pressure drops (up to 40 inches). Existing fan capacity is usually not sufficient to accommodate the added pressure drop. Therefore, booster fans or upgrades of the existing I.D. fans are typically required. FD fans can be affected as well by SNOX and NOXSO processes which involve combustion air streams.

Other existing components can also be affected. LIDS increases particulate loads at existing particulate collection equipment, SNRB requires a high-temperature baghouse with an integral SCR catalyst, and SNOX requires a high-efficiency baghouse to protect its catalyst and to function properly. This may require replacement of existing particulate collection equipment in retrofit applications. SNRB may also require air heater upgrading to realize the potential cost benefit of thermal performance improvement.

Air Toxics

The Clean Air Act Amendments of 1990 includes a list of 189 hazardous air pollutants, commonly known as air toxics, to be used in establishing emission regulations for various industries. Electric utility steam generating units were exempted from air toxics regulation until evaluations of potential health risks have been completed, but may then become subject to emissions standards. If these studies determine that power plant air toxics emissions are significant, regulations are to be promulgated by the year 2000.

Most of the listed substances are not applicable to power plants, but heavy metals associated with fossil fuels are a primary concern. Mercury is receiving special attention, but other heavy metals such as selenium and arsenic may also be included in air toxics regulations. Hydrochloric acid, formed from chlorides in coal during combustion, may be a concern as well.

Heavy metals can be present in flue gas in the form of fine (less than about 2 micron) particulates either alone or on the surface of flyash particles. The particulate forms are not readily controlled in wet FGD systems, because these systems do not effectively scrub fine particles. Instead, the particles pass through the absorber and are subsequently emitted. Removal of these particles requires improved, high-efficiency particulate collection systems.

Such systems are typically used in dry reagent-based FGD systems as an integral part of the process for removal of the reacted reagents. As a result, dry systems can be effective in removing particulate heavy metals. Wet systems generally do not include such systems
unless similar high-efficiency particulate collection systems exist ahead of the FGD system or unless wet particulate control devices are added downstream.

Heavy metals, particularly mercury and selenium, can also be present as elemental vapors or as oxide fumes (sub-micron sized particles). Heavy metals in these forms generally are not removed effectively by either wet or dry FGD systems. Collection can be accomplished by adsorption on activated carbon, and this removal method is more suited to use in combination with dry reagent-based FGD technologies.

Mercury also exists in the vapor state as other chemical species, notably mercuric chloride and methylmercury (an organometallic compound). Mercuric chloride is absorbed to varying degrees in both wet and dry reagent-based FGD systems. Because of this, the presence of chlorides aids in the removal of mercury by promoting the formation of mercuric chloride. Methylmercury is also absorbed to varying degrees in both wet and dry reagent-based FGD systems.

Hydrochloric acid can be present in flue gas in the vapor phase. Like acidic SO₂, this acid gas is readily removed by absorption in both wet and dry reagent-based FGD system. Collection efficiencies are high, and there is some evidence that hydrochloric acid collection can enhance SO₂ removal performance in dry systems by improving reagent particle surface characteristics. Wet FGD system SO₂ removals are not similarly enhanced. Instead, high chloride levels can impede limestone utilization for SO₂ absorption by providing a competing chemical reaction path.

**Illinois Coal Chlorides**

Chlorides in Illinois coals have a significant impact on FGD chemistry, design, and operation. Illinois coals from deep mines generally have chloride levels of 0.2 to 0.3%, which is higher than most other coals. These chlorides volatilize during combustion, forming hydrochloric acid in the flue gas.

In wet scrubbing processes, the chlorides are absorbed from the flue gas in the FGD absorber and accumulate in the recycle slurry. As water evaporates from the slurry, the chlorides concentrate to extremely corrosive levels and will attack carbon steel and some alloys. A blowdown stream is sometimes necessary to control chloride concentrations in the slurry to acceptably low levels.

If chloride levels in the slurry are below 8,000 ppm, construction of absorber and downstream wetted components with stainless steel metallurgy (such as grade 317LMN) is satisfactory. Levels above 8,000 ppm and up to 60,000 ppm require construction with higher alloys (such as alloy C276 or alloy C22) for corrosion resistance. When levels exceed 60,000 ppm, rubber lining may be necessary with liner replacement required at 5 to 10-year intervals.
Chloride concentrations in the slurry depend on coal chloride levels and on the amount of water leaving the system as a liquid. The effective water blowdown rate is in turn influenced by coal sulfur, sulfur removal efficiency, and the solids content of dewatered filter cake. Chloride levels above 60,000 ppm can easily be reached in LSFO systems with Illinois coals having high chloride contents. Chloride levels above 8,000 ppm are commonly reached in MgLime or LSDA systems, but levels above 60,000 ppm are rare in these systems because of the relatively high moisture content of the waste calcium sulfite solids.

High chloride levels can also affect process chemistry in wet systems. In LSFO applications, chlorides in the moisture of the gypsum solids can adversely affect gypsum quality and potential for sale. Absorption of chlorides affects MgLime systems by increasing the magnesium content required in the circulating slurry. In LSDA systems, additional soda ash must be used to neutralize absorbed chlorides in order to maintain the desired alkalinity of the solution.

In dry scrubbing processes, chlorides react with lime and over 90% are removed as calcium chloride solids. Because calcium chloride is hygroscopic, it promotes SO₂ absorption on lime particle surfaces and enhances SO₂ removal in the dry scrubber. Thus chlorides are beneficial to system performance. Although dry systems do not encounter the corrosive conditions experienced in wet FGD systems, downstream ductwork and electrostatic precipitator internals are subjected to more severe service when firing coals high in chlorides. Protective coatings are generally needed.

Catalytic or regenerable sorbent processes are generally not affected by chlorides found in Illinois coal applications, because the chlorides remain in the flue gas.

**Alternative Emissions Control Strategies**

Existing utilities firing high sulfur coals and faced with controlling emissions under Phase II have several alternative strategies available:

- Switch to firing low-sulfur Western coal
- Continue to fire Illinois coal and purchase SO₂ allowances, or
- Continue to fire Illinois coal and retrofit an FGD system.

These alternatives were examined for an existing 500 MW power plant. A 0.48% sulfur Wyoming coal was assumed for coal switching. Use of this coal instead of 3% sulfur Illinois coal would provide a 76% reduction in SO₂ emissions. Conversion to this coal would involve capital costs associated with handling and firing the western coal.

Purchase of SO₂ allowances equivalent to the same 76% reduction was considered as a second alternative at a price of $150 per ton and also at an assumed future price of $300 per ton.
Retrofitting a Limestone Forced Oxidation FGD system was examined for comparison of FGD installation to these other alternative strategies. This comparison used a capital cost of $128/kW, the value estimated for new installations. While aggressive, costs for retrofit projects can approach this level particularly if recent advances are adopted. Since LSFO can provide a 95% SO₂ removal, the additional removals beyond 76% were assumed to be valued at $150 and $300 per ton as an operating credit.

Case Study

The technical and cost data developed in this study was applied to an actual Midwestern utility who is affected by Phase II of the Clean air Act Amendments. The analysis was done to show the usefulness of information developed in this study to anyone evaluating FGD options. All the technologies were applied to the utility FGD evaluation and a Limestone Forced Oxidation system was the recommended technology. LSFO was recommended because it was found to be the low cost, low risk option.

This particular utility had an existing LSFO system for another unit on site and the existing FGD system had considerable installed spare capacity in the reagent preparation and waste dewatering subsystems. Utilizing this previously installed capacity allowed the capital and levelized cost of the LSFO system to be reduced by approximately 33%. This illustrates that a utility with an existing FGD system is a particularly attractive site for a second FGD system.

CONCLUSIONS AND RECOMMENDATIONS

1. The estimated rates of chemical consumptions, utility usages, waste generations, and byproduct productions for the nine FGD technologies surveyed are summarized in Table 1.

2. The estimated capital requirements, O&M costs, credits for byproduct sales and heat recovery, and total levelized FGD costs for the nine FGD technologies surveyed are summarized in Table 2. Unit prices used in the economic analysis are listed in Table 3. These economic evaluations showed that for a typical 500 MW unit using 3% sulfur Illinois coal, the impact on the cost of electricity would range from 6.6 to 8.8 mil/kWh for all technologies except SNRB. Levelized costs for SNRB were penalized by the high lime consumption required to achieve 95% SO₂ removal. At lower removals, costs for SNRB would be more competitive with other technologies.

3. Capital requirements for the nine FGD technologies surveyed are illustrated and compared in Figure 10. The technologies which treated only SO₂ had capital costs ranging from $111 to $148 per kW, while the technologies which treated both SO₂ and NOₓ had capital costs ranging from $256 to $379 per kW.
4. Levelized costs for the nine FGD technologies surveyed are illustrated and compared in Figure 11. These costs were slightly higher for technologies which removed NOx than for technologies which did not.

5. The sensitivity of levelized FGD costs to power plant size is shown in Figure 12. All technologies were affected by plant MW size, reflecting capital cost economies of scale.

6. The sensitivity of levelized FGD costs to coal sulfur content is shown in Figure 13. Costs for most technologies rose as sulfur levels were increased, but costs for SNOX, NOXSO, and Copper Oxide decreased instead because of the additional credits gained by increasing production of byproduct sulfuric acid.

7. A comparison of FGD to coal switching and to purchasing SO2 allowances as alternative control strategies for existing units is shown in Figure 14. This plot indicates economics currently favor conversion to western coal over retrofitting a LSFO FGD system for coal price premiums less than about $0.30 to $0.40 per MBtu. Purchasing allowances is also less expensive than LSFO unless allowance prices rise to $300 per ton. These conclusions would not apply for new units.

8. The economic results of the case study are shown in Figure 15 (Capital Costs) and Figure 16 (Levelized Costs). As shown by these figures, LSFO using existing reagent and waste handling equipment was the lowest cost alternative for this application with a $173/kW capital cost and a 5.03 mil/kWh levelized cost.
DISCLAIMER STATEMENTS

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# Table 1

**FGD Operating Chemicals and Utilities**

500 MW New Unit

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<thead>
<tr>
<th></th>
<th>Chemicals Used</th>
<th>Utilities</th>
<th>Other</th>
<th>Waste Generation</th>
<th>Byproduct</th>
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<tbody>
<tr>
<td></td>
<td>limestone (lb/hr)</td>
<td>lime (lb/hr)</td>
<td>soda ash (lb/hr)</td>
<td>calcium chloride (lb/hr)</td>
<td>ammonia (lb/hr)</td>
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<tr>
<td>LSFO</td>
<td>38,092</td>
<td></td>
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<tr>
<td>MagLime</td>
<td>22,291</td>
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<td>LSAD</td>
<td>38,087</td>
<td>2,311</td>
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<tr>
<td>LIDS</td>
<td>65,400</td>
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</tr>
<tr>
<td>GSA</td>
<td>28,800</td>
<td>1,297</td>
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<td></td>
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</tr>
<tr>
<td>SNRB</td>
<td>53,400</td>
<td></td>
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<td></td>
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<tr>
<td>SNOX</td>
<td>991</td>
<td>1,076</td>
<td></td>
<td></td>
<td></td>
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<tr>
<td>NOXSO</td>
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<tr>
<td>CuO</td>
<td>991</td>
<td>6,040</td>
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Estimates for 500 MW, 3% sulfur coal, 95% SO2 removal, and 65% capacity factor
<table>
<thead>
<tr>
<th></th>
<th>Total Capital Requirement $1,000</th>
<th>$/kW</th>
<th>Operating &amp; Maintenance Costs</th>
<th>Variable O&amp;M</th>
<th>Credits</th>
<th>Total Levelized FGD Cost</th>
<th>FGD Acid (mil/kWh)</th>
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<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Fixed O&amp;M</td>
<td></td>
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</tr>
<tr>
<td></td>
<td>Operating Labor $1,000/yr</td>
<td></td>
<td>Mice Material $1,000/yr</td>
<td>Mice Labor $1,000/yr</td>
<td>Admin Labor $1,000/yr</td>
<td>Chemicals $1,000/yr</td>
<td>Sorbent $1,000/yr</td>
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<tr>
<td>LSFO</td>
<td>63,817</td>
<td>128</td>
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<td>1,902</td>
<td>1,268</td>
<td>819</td>
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<td>MagLime</td>
<td>55,556</td>
<td>111</td>
<td>1,617</td>
<td>1,641</td>
<td>1,094</td>
<td>813</td>
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<td>LSDA</td>
<td>71,775</td>
<td>144</td>
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<td>1,741</td>
<td>1,161</td>
<td>787</td>
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<td>LIDS</td>
<td>70,650</td>
<td>141</td>
<td>1,248</td>
<td>1,905</td>
<td>1,056</td>
<td>692</td>
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<tr>
<td>GSA</td>
<td>73,650</td>
<td>148</td>
<td>632</td>
<td>1,154</td>
<td>770</td>
<td>481</td>
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<td>SNRB</td>
<td>142,016</td>
<td>275</td>
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<tr>
<td>SNOX</td>
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<td>379</td>
<td>624</td>
<td>2,958</td>
<td>1,972</td>
<td>779</td>
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<td>NOXSO</td>
<td>143,835</td>
<td>279</td>
<td>832</td>
<td>1,878</td>
<td>1,252</td>
<td>625</td>
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<tr>
<td>CuO</td>
<td>131,378</td>
<td>256</td>
<td>832</td>
<td>1,379</td>
<td>920</td>
<td>526</td>
<td>494</td>
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Estimates for 500 MW, 3% sulfur coal, 95% SO₂ removal, and 65% capacity factor
Constant Dollar Basis
Table 3
FGD Operating Labor, Chemicals, and Utilities
Unit Costs for Economic Analysis

**Reagents & Chemicals**

<table>
<thead>
<tr>
<th>Chemical</th>
<th>Cost</th>
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<tr>
<td>Limestone</td>
<td>$15/ton</td>
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<tr>
<td>Lime</td>
<td>$65/ton</td>
</tr>
<tr>
<td>Soda ash</td>
<td>$93/ton</td>
</tr>
<tr>
<td>Calcium chloride</td>
<td>$200/ton</td>
</tr>
<tr>
<td>Ammonia</td>
<td>$175/ton</td>
</tr>
<tr>
<td>Sorbent</td>
<td>$1.50/lb  (NOXSO)</td>
</tr>
<tr>
<td></td>
<td>$2.00/lb  (Copper Oxide)</td>
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**Utilities**

<table>
<thead>
<tr>
<th>Utility</th>
<th>Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Power usage</td>
<td>$0.050/kWh</td>
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<tr>
<td>Heat recovery</td>
<td>$0.020/kWh</td>
</tr>
<tr>
<td>Fresh water</td>
<td>$0.60/1,000 gal</td>
</tr>
<tr>
<td>Low pressure steam</td>
<td>$3.00/1,000 lb</td>
</tr>
<tr>
<td>Natural gas</td>
<td>$2.50/MMBtu</td>
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**Wastes and Byproducts**

<table>
<thead>
<tr>
<th>Wastes</th>
<th>Cost</th>
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<tbody>
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<td>Sludge disposal</td>
<td>$10/ton</td>
</tr>
<tr>
<td>Fixative lime</td>
<td>$65/ton</td>
</tr>
<tr>
<td>Sulfuric acid</td>
<td>$52.50/ton (100% basis)</td>
</tr>
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**Labor**

<table>
<thead>
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<th>Labor</th>
<th>Cost</th>
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<tbody>
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Figure 3
Limestone Dual Alkali

Figure 4
Limestone Injection with Dry Scrubbing
Figure 5
Gas Suspension Absorption

Figure 6
SNRB Process
Figure 9
Copper Oxide Process

SO₂ absorbed on moving bed of copper oxide in absorber

Flue gas from economizer

Ammonia added for NOₓ removal

Ammonia

Moving Bed Absorber

Spent sorbent

SO₂ desorbed from sorbent by treatment with reducing gas

Sorbent recycled to absorber (copper in metallic form)

Reform

Reducing gas mixture of carbon monoxide and hydrogen produced from methane and steam in reformer

Water

Natural gas

Air

Regenerator

Byproduct elemental sulfur or sulfuric acid

Sulfur recovery unit

Clean plant or contact sulfuric acid plant

Contains SO₂ removed from flue gas

Fly ash to disposal

Spent sorbent

Desulfurized flue gas goes to air heater

Vent to flue gas
Figure 11
Total FGD Levelized Cost - 500 MW New Unit
Summary Comparison
Figure 12
FGD Levelized Cost Sensitivity to Unit Size
Summary Comparison

Levelized Cost (MILL/kWh)

100 MW Unit
300 MW Unit
500 MW Unit

CuO
NOXSO
SNOX
SNRB
GSA
LIDS
LSDA
MagLime
LSFO
Figure 13
FGD Levelized Cost Sensitivity to Coal Sulfur
Summary Comparison
Figure 14
Alternative SO₂ Emission Control Strategies
Cost Comparison - 500 MW Unit

Levelized Cost (mill/kWh)

Incremental Cost of Western Coal ($ per MMBtu)

- Conversion to Western Coal
- Limestone Forced Oxidation FGD
- Purchase Emission Allowances

FGD Installed at $128/kW and SO₂ Allowances Valued at $150/ton
FGD Installed at $128/kW and SO₂ Allowances Valued at $300/ton

SO₂ Allowances Purchased at $300/ton
SO₂ Allowances Purchased at $150/ton

Conversion to Western Coal at $110/kW
Conversion to Western Coal at $50/kW
Figure 15
Midwestern Unit Case Study
Capital Cost for Alternative FGD Technologies

![Capital Cost Chart]

- Conv. LSFO: 251
- Adv. LSFO: 173
- Mg Lime: 237
- Dual Alkali: 283
- Amm. Scrub.: 364
- GSA: 234
- SNOX: 470
- NOXSO: 370
- CuO: 336
- SNRB: 364
Figure 16
Midwestern Unit Case Study

Levelized Cost for Alternative FGD Technologies

Annual Capacity Factor - 66%

Levelized Cost (MILL/KWH)
PROJECT MANAGEMENT REPORT
June 1 through August 31, 1995

Project Title: SURVEY OF FLUE GAS DESULFURIZATION SYSTEMS FOR ILLINOIS COAL

ICCI Project Number: 94-1/2.1B-2
Principal Investigator: Joseph N. Darguzas, Sargent & Lundy
Other Investigators: David G. Sloat, Sargent & Lundy
Raj Gaikwad, Sargent & Lundy
Jeri Penrose, Sargent & Lundy
Project Manager: Franklin I. Honea, Illinois Clean Coal Institute

COMMENTS

Tasks 1A, B, and C; 2A and B; 3A; 4; and 5 are complete.

At the request of the ICCI project manager, we added, at no extra charge, one FGD technology that was not in our original scope. This new technology is gas suspension adsorption (GSA). This technology was recently described by DOE to have significant cost and air toxic benefits compared to conventional limestone FGD.

We also included, at no extra charge, one case study in our conclusions. This case study was for a midwestern utility station and we used our cost data to analyze the cost of FGD at this station.

We will be sending a final report to ICCI which includes a detail writeup of each process as well as the summary final report.
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<tr>
<th>Quarters *</th>
<th>Types of Costs</th>
<th>Direct Labor</th>
<th>Fringe Benefits</th>
<th>Materials &amp; Supplies</th>
<th>Travel</th>
<th>Major Equipment</th>
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* - Cumulative by Quarter
SURVEY OF FLUE GAS DESULFURIZATION SYSTEMS FOR ILLINOIS COAL

Cost By Quarters

- Projected Expenditures $78,050
- Estimated Projected Expenditures $78,050

Total Illinois Clean Coal Institute Award $78,050
### Schedule of Project Milestones

<table>
<thead>
<tr>
<th>Task</th>
<th>Description</th>
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<td>1A</td>
<td>Process description for each technology</td>
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<td>1B</td>
<td>Process flow diagrams</td>
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<td>1C</td>
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<td>ICCI conference presentation</td>
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<td>Progress reports</td>
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**Task 1** Survey of 9 FGD technologies

**Subtask 1A** Process description for each technology

**Subtask 1B** Process flow diagrams

**Subtask 1C** Design basis

**Task 2** Cost estimates

**Subtask 2A** Capital Costs

**Subtask 2B** Operating costs

**Task 3** Final report

**Subtask 3A** Draft to ICCI

**Subtask 3B** Issue report

**Task 4** Summary PAPER

**Task 5** ICCI conference presentation

**Task 6** Progress reports