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FLUE-GAS DESULFURIZATION: REVIEW OF SELECTED COMMERCIAL AND ADVANCED TECHNOLOGIES

by

Jerry L. Gillette and Shen-Yann Chiu

FOSSIL ENERGY PROGRAM

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Energy and Environmental Systems Division
Integrated Assessments and Policy Evaluation Group

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SUMMARY

Several advanced flue-gas-desulfurization (FGD) technologies are evaluated with respect to their current status, estimated costs, and anticipated SO_2 removal efficiencies. The basic lime and limestone wet scrubbing processes are also evaluated to provide a base case against which the advanced technologies can be compared. A brief overview of each technology covered in the report is presented below.

FGD TECHNOLOGIES

- Wet Scrubbing with Lime or Limestone

The lime and limestone FGD processes are considered as one reference technology, or base case, for the advanced technologies discussed in this report. This method of SO_2 removal has been applied in many coal-fired electric-generating stations, and many more units are in the design or construction phase. Historically, capital investments and operating costs vary greatly from application to application. The different costs reflect significantly different site factors, the sulfur content of the coal used, different SO_2 removal requirements, etc. Major advantages of this base-case technology include the extensive experience gained with it to date and the availability of the materials needed. Disadvantages include a high rate of forced outages, corrosion and erosion problems, and the need to dispose of great quantities of sludge.

- Dry Injection/Baghouse Filtering

The dry injection/baghouse filtering technology is a comparatively simple mechanical system with a comparatively low capital investment. Low operating and maintenance costs and a low rate of forced outages are also characteristics of this technology. Because it is a dry FGD process, the water requirements are very much less than the base-case lime/limestone technologies. One disadvantage of this system is a somewhat low SO_2 removal efficiency, which may limit its application to low-sulfur coals. Additional potential problems include dependence on a limited supply of nahcolite and the need to dispose of a relatively soluble sodium-based salt, which could leach into underground water supplies. This FGD technology is in the testing stage, with prototypes at 60-MW and 80-MW stations.

- Spray Drying/Baghouse Filtering

The spray drying/baghouse filtering technology represents an improvement over the dry injection/baghouse filtering process in that sorbents other than the scarce nahcolite can be used; a somewhat higher SO_2 removal efficiency can be achieved; and, depending upon the sorbent used, the waste-disposal difficulties can be significantly reduced. In principle, this technology can be used with all types of coal; economic considerations, however, may limit its application to coals with a sulfur content of less than 2.5-3%. The technology has been commercially applied in both utility and industrial facilities.

- Aqueous-Carbonate Dry Scrubbing

The aqueous-carbonate process combines the spray-dryer technology with methods of regenerating the sorbent material and producing a marketable end product with the sulfur removed from the flue gas. This FGD technology can therefore greatly reduce the amount of sorbent (Na_2CO_3) required and can also produce revenues that partially offset its own costs. Regenerable and sulfur-recovery techniques make the system complex, however, and this complexity raises both investment costs and operating and maintenance costs. On the other hand, SO_2 removal efficiencies can potentially exceed those that are possible with other advanced FGD technologies. A 100-MW test facility is currently under construction.

- Copper-Oxide Dry Scrubbing

The dry copper-oxide process offers the potential of significant NO_x removal as well as very good SO_2 removal. It is a regenerable process that produces a marketable end product. Capital expenditures for this process are expected to be very high. The energy required to regenerate the reactor beds so that they will accept additional SO_2 is estimated to be a significant fraction of the powerplant output. This FGD process is currently in the pilot-scale testing phase.

- Bergbau-Forschung Dry Scrubbing

The Bergbau-Forschung activated-char process has been tested and developed to combine high levels of SO_2 and NO_x removal with sorbent regeneration and sulfur recovery. Although the complexity of the system might be expected to result in high capital investments, the economic estimates to date show a wide range of values. Removal efficiencies for SO_2 are expected to be as high as 95%, even for high-sulfur coals. This process is currently being tested on units ranging from 20 to 50 MW. Systems for still larger units have been designed in which the principal parameters of the adsorption section were kept constant and modular contruction was utilized in the desorption section.

- Sodium-Sulfite-Based Double-Alkali Wet Scrubbing

Double-alkali scrubbing is a wet, regenerable process combining absorption of SO_2 with aqueous alkali, and regeneration of the absorbent with lime or limestone. The double-alkali systems utilize a clear sodium-sulfite-based absorption solution. They reduce the problems of plugging, scaling, and erosion. Existing systems remove SO_2 with 90-95% efficiency. While some systems have had mechanical or chemical problems, they have shown themselves reliable; less than 10% of their total operating time has been interrupted with forced outages. This FGD technology has good retrofitting potential based on the small size of its components. The process requires a large land area for disposing of the solid waste it generates. Economically, double-alkali systems appear to be competitive with the wet lime and limestone systems. The process has been commercially applied in the U.S. Three full-scale demonstration systems are operating with coal-fired utility boilers, and several commercial units are in operation with coal- and oil-fired industrial boilers. Further development work is needed to evaluate, characterize, and compare full-size coal-fired demonstration facilities; to test systems using limestone as a regenerant; and to develop methods for upgrading the quality of sludge and strategies to control the multimedia environmental effects of effluents and emissions.

- DOWA Double-Alkali Wet Scrubbing

The DOWA double-alkali FGD process (also called the DOWA aluminum-sulfate-limestone process) uses a basic aluminum-sulfate solution as the absorbent and limestone as the precipitant to remove SO_2 from flue gases. This process was developed by DOWA Mining Co. in Japan. It has been commercially applied in Japan to desulfurize flue gas from an oil-fired boiler and waste gases from smelters, roasters, and sulfuric-acid plants. Its applicability to coal-fired boilers is now being tested in the U.S. The SO_2 removal efficiency of the DOWA process appears to be greater than 90%. The SO_2 removed in the process is converted into gypsum, which can either be sold or disposed of on land. The process is simple and easy to operate. Capital and operating costs, based on a preliminary estimate, are competitive with those of the lime and limestone processes. More study is needed to evaluate process design parameters, optimize process performance, and determine the environmental effects. Furthermore, economic evaluation of the process needs to be updated.

- Chiyoda-121 Wet Scrubbing

The Chiyoda FGD process identified as Thoroughbred 121 (or CT-121) is an improved version of the conventional lime and limestone wet-scrubbing processes. It uses a compact reactor that combines in one vessel sulfur-dioxide absorption, sulfite oxidation, acid neutralization, and gypsum crystallization. Combining all these processes could greatly simplify the design and operation of an FGD system. Development of this FGD process is relatively recent. Results from limited pilot and prototype tests indicate a high degree of reliability, efficient desulfurization, and ease of operation. Preliminary cost estimates also reveal that this FGD process could be economically competitive with lime and limestone scrubbing. A full-scale demonstration by a utility company appears to be a logical next step before commercialization.

- Magnesia Wet Scrubbing

The magnesia-slurry scrubbing technology is a regenerable, recoverable FGD process utilizing magnesium hydroxide to absorb SO_2 in a wet scrubber. The aqueous slurry of magnesium sulfite, formed in the scrubber, is dried and calcined to regenerate magnesium oxide. The magnesium oxide is hydrolized and reused in the absorber. The SO_2-rich stream is sent for sulfur recovery. The system includes relatively complicated chemical processes; however, the absorbent can be regenerated offsite to reduce operating complexity and costs. The MgO process has a 90% or greater SO_2 removal efficiency. It is relatively energy-intensive, requiring a great deal of fuel for magnesium-oxide regeneration. The magnesium-slurry FGD process is being commercially demonstrated in the U.S. on both oil-fired and coal-fired boilers. Low system availability has been a problem. Future development should also focus on improvement of system reliability and use of coal for absorbent regeneration. The latter would reduce system dependence on the nation's limited supplies of oil and gas.

- Wellman-Lord Wet Scrubbing

Wellman-Lord is an aqueous process that employs a sodium-sulfite scrubbing solution to remove SO_2 from flue gas. Thermal regeneration enables the system to recover the sulfite and produce a concentrated stream of SO_2. This process has been applied commercially both in the U.S. and overseas to desulfurize flue and waste gases from oil- and coal-fired boilers, nonferrous smelters, sulfuric-acid plants, and Claus plants. This FGD process has all of the advantages associated with sodium-sulfite-based scrubbing: a high SO_2 removal efficiency, no plugging or scaling in scrubbing, and a low liquid-to-gas ratio. It is a closed-loop operation, producing marketable end products with no large-scale solid-waste disposal problems. The regeneration loop is a complicated process requiring a relatively high energy input and relatively higher capital and operating costs, therefore, than throwaway processes. Further development is needed to investigate specific process improvements; to evaluate the process performance in full-scale demonstration with coal-fired boilers; and to test the Wellman-Lord system in combination with downstream sulfur-reduction systems, specifically those using coal as a reducing agent.

FGD-RELATED TECHNOLOGIES

- RESOX Sulfur-Recovery Process

The RESOX process is one of several that have been developed to recover sulfur from SO_2-rich streams. RESOX uses coal as a direct reductant for converting sulfur dioxide to elemental sulfur, a product that appears to be more desirable in several respects than such other compounds as sulfuric acid and ammonium sulfate. The process involves relatively simple unit operations, and it could be one of the less expensive sulfur-recovery processes. However, the development of this process is not far enough along for definitive cost evaluation. Development of the RESOX process was initiated in the early 1970s by Foster Wheeler Corp. Present efforts by Foster Wheeler to move the process toward full commercialization include pilot tests and a prototype demonstration. Further system studies are needed to test its adaptability to various front-end FGD systems (e.g., Wellman-Lord) and the feasibility of using bituminous coal as a direct reductant -- as well as to improve SO_2 conversion efficiency, to demonstrate the usages of residual coal (e.g., as a fuel or SO_2 absorbent), and to improve the operation and design of sulfur-condensing and sulfur-purification systems.

- Direct Combustion of Limestone

Two SO_2-control processes currently under investigation involve the direct combustion of limestone. One of these processes utilizes coal/limestone pellets as the boiler fuel. The other process uses a pulverized coal/limestone mixture in a low-temperature burner. In each case the sulfur released upon burning the coal is tied up as calcium sulfite or sulfate before the flue gas leaves the burner. These processes are both in the early stages of development.

TABULAR OVERVIEW

Selected economic, environmental, and technical data about the foregoing FGD processes are summarized in Table 1. Cost estimates for each FGD technology are given in terms of unit-capacity requirements. These data are mostly engineering estimates based on varying assumptions. They should not be understood as hard data. The energy requirements are expressed in terms of percent of energy input to an uncontrolled power plant having the same net electrical output. Sulfur-dioxide removal efficiencies represent best estimates of the long-range capability of the process. The sorbent-regeneration and sulfur-recovery columns indicate the current state of the art of the process. In almost all cases, the potential exists for sorbent regeneration or sulfur recovery or both. Taking advantage of that potential would of course increase system complexity and investment costs and energy requirements. The development-status column indicates where each of the processes stands as of Feb. 1981.

Table 1 Economic/Environmental/Technical Overview of Selected FGD Technologies

Process[a]	Capital Investment[b] ($/kW, 1980 Dollars)	O & M Costs[c] (mill/kWh, 1980 mills)	Energy Required (%)[d]	SO_2 Removal Efficiency (%)	Sorbent Regeneration	Sulfur Recovery	Present Stage of Development[e]
Lime wet scrubbing	140 to 168	3.2 to 5.2	4 to 5	>90	No	No	Commercial
Limestone wet scrubbing	147 to 178	2.7 to 5	2.5 to 4	80	No	No	Commercial
Dry injection/baghouse filtering	50 to 90	2 to 5	<1	70 to 80	No	No	Prototype
Spray drying/baghouse filtering	90 to 120	1 to 3	<1	90	No	No	Commercial
Aqueous-carbonate dry scrubbing	173 to 274	4.6 to 5.4	6	90 to 95	Yes	Yes	Demonstration
Copper-oxide dry scrubbing	160 to 200	3 to 5	10 to 11	90[f]	Yes	Yes	Pilot testing
Bergbau-Forschung dry scrubbing	170 to 210	4.4 to 5.9	6 to 7	80 to 95[f]	Yes	Yes	Prototype
Sodium-sulfite-based double-alkali wet scrubbing	150 to 180	1.9 to 5	2.5 to 4	>90	Yes	No	Demonstration and commercial
DOWA double-alkali wet scrubbing	150 to 180	2 to 5	2 to 3	>90	Yes	No[g]	Prototype and commercial
Chiyoda-121 wet scrubbing	h	h	2 to 3	85 to 90	No	No[g]	Prototype and commercial
Magnesia wet scrubbing	160 to 190	2.5 to 5.5	5 to 10	>90	Yes	Yes	Demonstration
Wellman-Lord wet scrubbing	155 to 190	2.5 to 5.5	12 to 25	>90	Yes	Yes	Demonstration and commercial

[a] This tabularized list of processes differs slightly from the way they are treated in the body of the report (Secs. 2 through 14), where lime scrubbing and limestone scrubbing are treated as one technology with two variations. This tabularized list also differs from the way the FGD technologies are summarized in the text immediately preceding Table 1. The text overview includes two technologies (direct combustion of limestone and the RESOX sulfur-recovery process) that, strictly speaking, do not remove SO_2 from flue gas. Rather, the RESOX process may be used as a followup to those FGD processes that allow sulfur recovery; and limestone combustion is a means of preventing or reducing the formation of sulfur dioxide in flue gas before the gas is emitted.

[b] Includes direct and indirect investment, land and working capital, and the assumed allowance of a spare module.

[c] Operating and maintenance costs include raw material, operating labor and supervision, utilities, maintenance, and solid-waste disposal.

[d] Percent of the energy input to an uncontrolled power plant of the same net electrical output.

[e] Relative to coal-fired boiler applications.

[f] NO_x removal is also a significant capability.

[g] Gypsum, however, is produced and can be sold as a by-product; or it can be thrown away.

[h] Cost estimates were not available when the data in this report were gathered.

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ABSTRACT

Selected technologies for the desulfurization of flue gas are reviewed and evaluated. One of the technologies, wet scrubbing of the gas with a slurry containing either lime or limestone, has attained widespread commercial usage and for this reason is presented as the base case against which 10 second-generation systems of flue-gas desulfurization (FGD) are compared. The new technologies include chemical-treatment, or scrubbing, processes and processes that combine scrubbing and filtering of the flue gas. The two combination processes use baghouse filters, with the SO_2-collecting material either injected dry or sprayed into the system. The eight straight-scrubbing processes use either wet or dry collecting materials and are identified as follows: aqueous-carbonate, copper-oxide, Bergbau-Forschung, sodium-sulfite-based double-alkali, DOWA double-alkali, Chiyoda-121, magnesia, and Wellman-Lord. In addition to reviewing those 10 processes in detail, the report includes information on two new technologies that are related to FGD. One of them combines limestone with the coal being burned as boiler fuel -- the purpose being to reduce the formation of sulfur in flue gas before the gas is emitted. The other technology is a sulfur-recovery process, identified as RESOX, and is designed as a followup to those FGD technologies proper that allow sulfur removed from the flue gas to be recovered.

1 INTRODUCTION

Present national energy policy calls for increased production and consumption of abundant domestic coal. This emphasis on increased coal utilization and a related emphasis on tighter environmental-control requirements have prompted development of numerous air-cleanup methods. One of these cleanup methods is flue-gas desulfurization (FGD), the generic term for processes that chemically or physically reduce the sulfur-dioxide level in the exhausts of facilities burning fossil fuels. More than 30 separate technologies have been developed to accomplish this basic objective.

Flue-gas desulfurization is still in its infancy. Most of the commercial applications in the U.S. have been limited to the first-generation FGD technology, lime/limestone wet scrubbing. Existing lime/limestone systems have had certain problems: a high incidence of forced outages, corrosion and scaling of process components, and, in particular, problems relating to containment and disposal of the large quantities of wet sludge generated by the process. Specific improvements and modifications of the conventional lime/limestone scrubbing process are being studied to provide technical solutions to these problems.

In addition to conventional lime/limestone scrubbing, several of the advanced (i.e., second-generation) FGD technologies give promise of improved system reliability (fewer forced outages), less energy consumption, lower capital investment and operating costs, and less severe waste-disposal problems. These technologies are in various stages of development: many have been tested at the bench and in small-scale industrial applications, some have been operated in Japan on oil-fired boilers, and some are being tested as prototype and demonstration systems with fossil-fueled equipment in the U.S.

Objective. This report -- prepared as part of an Argonne National Laboratory (ANL) effort (Task #49602, Systems Assessment of Fossil Fuel Utilization Technologies) -- evaluates the current state of selected FGD technologies, with emphasis on advanced technologies. It presents an economic/environmental/technical overview of the technologies. It also addresses issues pertinent to the research and developmental planning of the advanced technologies considered.

Scope. Most of the 10 second-generation FGD technologies evaluated in this report are outlined in the Department of Energy (DOE) report entitled "Multiyear Program Plan for Advanced Environmental Control Technology" (June 27, 1979). They include chemical-treatment, or scrubbing, processes and processes that combine scrubbing and filtering. Each scrubbing process is classified wet or dry depending on whether its SO_2-collecting material (sorbent) is in a liquid or nonliquid state -- and also classified according to whether the sorbent is regenerable and whether sulfur removed from the flue gas can be recovered. The 10 second-generation FGD technologies covered in this report are:

Advanced Lime-Limestone Wet-Scrubbing Systems (Sulfur Nonrecoverable)

- Sodium-sulfite-based double-alkali scrubbing (sorbent regenerable)
- DOWA double-alkali scrubbing (sorbent regenerable)
- Chiyoda-121 scrubbing (sorbent nonregenerable)

Other Advanced Wet-Scrubbing Systems (Recoverable/Regenerable)

- Magnesia scrubbing
- Wellman-Lord scrubbing

Advanced Dry-Scrubbing/Filtering Systems (Nonrecoverable/Nonregenerable)

- Dry injection/baghouse filtering
- Spray drying/baghouse filtering

Advanced Dry-Scrubbing Systems (Recoverable/Regenerable)

- Aqueous-carbonate scrubbing
- Copper-oxide scrubbing
- Bergbau-Forschung, or activated-char, scrubbing

In addition to these desulfurization technologies, the RESOX process, a new method using coal as a reductant to recover sulfur from the SO_2-rich stream of recoverable FGD systems, is reviewed. Also reviewed is a new technology for burning limestone with coal as boiler fuel to limit formation of SO_2 in flue gases.

To provide for a basis of comparison, the conventional lime/limestone scrubbing is also presented.

Major topics addressed for each FGD technology include:

- Process description and chemistry
- Other technical aspects of the process
- Resource requirements (water, chemical, power)
- Environmental implications
- Cost estimates
- Present stage of development
- Research and development programs underway, if any

The technologies described in this document apply mainly to utility applications. Industrial applications are noted where they are appropriate.

Approach. We thoroughly reviewed the available literature for each FGD technology considered. In the limited number of discussions we also held, with those in the private and public sectors whose organizations are doing developmental work on the technologies, the subjects discussed included: specific FGD problems, ongoing activities, technical judgments, and program plans and needs. (More such discussions will probably be necessary if this evaluation continues into a second, more detailed phase.) No effort was made to compare the technologies on a common design and cost basis.

Most of the advanced FGD processes are still in various testing stages and have not yet attained commercial status. As a result, there is significant uncertainty in some of the performance parameters -- particularly capital investment, operating cost, and energy requirement. In some cases, a testing program has thus far concentrated on a limited range of coal types. This in turn limits extrapolations that can be made to other situations. It is uncertain, for example, whether tests using low-sulfur western coal are applicable in situations where medium- or high-sulfur midwestern coal would be burned. A similar situation exists with respect to performance and cost parameters of the basic lime and limestone processes. Those parameters show great variations that reflect different coal types, different SO_2 removal efficiencies, and different local-site conditions.

Despite the uncertainty just mentioned, however, the authors believe the information in this report presents an accurate account of the principal economic, environmental, and technical aspects of the advanced FGD technologies selected for review.

2 LIME AND LIMESTONE WET SCRUBBING: BASE-CASE TECHNOLOGY

SUMMARY

The lime and limestone FGD processes are considered as one reference technology, or base case, for the advanced technologies discussed in this report. This method of SO_2 removal has been applied in many coal-fired electric-generating stations, and many more units are in the design or construction phase. Historically, capital investments and operating costs vary greatly from application to application. The different costs reflect significantly different site factors, the sulfur content of the coal used, different SO_2 removal requirements, etc. Major advantages of this base-case technology include the extensive experience gained with it to date and the availability of the materials needed. Disadvantages include a high rate of forced outages, corrosion and erosion problems and the need to dispose of great quantities of sludge.

2.1 TECHNICAL ASPECTS

The technical performance of the lime/limestone base case technology is presented below.

2.1.1 Description of Processes

The bulk of the FGD experience to date has been derived from conventional wet lime and limestone processes. Of the 18,504 MW of FGD capacity operational in the United States as of June 1979, the wet lime or limestone process accounted for approximately 90%.[1] In the near-term future, strong reliance on this FGD technology can be expected to continue: about 83% of the 17,135 MW of FGD capacity currently under construction is based on the lime/limestone technology.

The lime and the limestone processes are basically very similar, but some differences do exist. The principal difference in the two processes is in the slurry-preparation phase. In the limestone process, the limestone is wet-ground and then pumped to a slurry feed tank adjacent to the scrubbers. This makeup slurry and recycled clear liquor from the disposal pond are pumped to the absorber-effluent hold tank, where it is blended with the partially reacted slurry draining from the absorber. Slurry is recirculated between the effluent hold tank and the absorber, where the gas is scrubbed. Inside the scrubber, the slurry contains 8-17% solids. A bleed stream of reaction products is pumped from the hold tank to the waste-disposal system or to a settling pond. The flue gas exiting the scrubber flows through a slurry-entrainment separator and is then reheated and exhausted from the stack. A simplified flow diagram for a limestone-slurry FGD process is shown in Fig. 2.1.

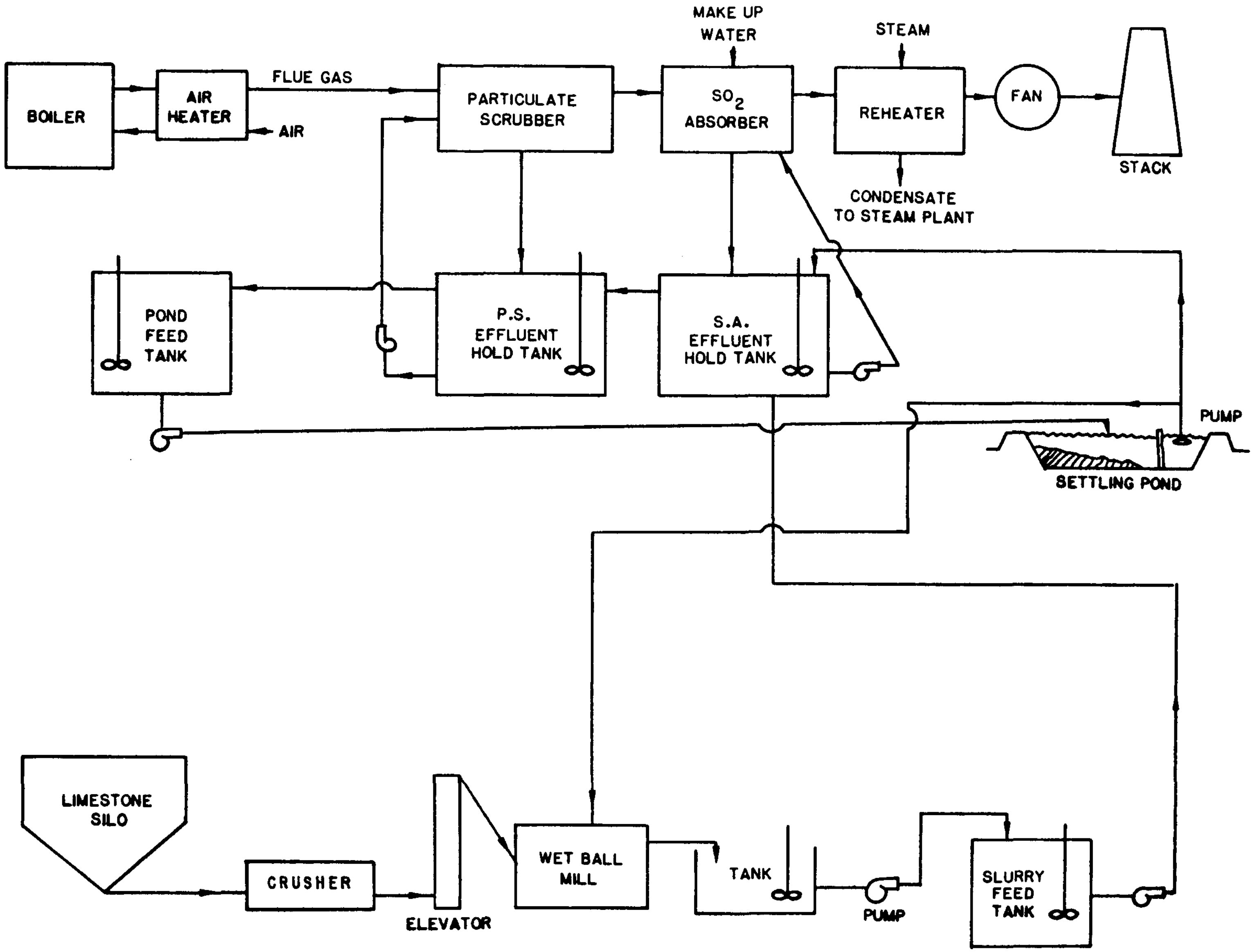


Fig. 2.1 Flow Diagram of Limestone Wet-Scrubbing FGD Process

The limestone-slurry process in Fig. 2.1 and the lime-slurry process in Fig. 2.2 are quite similar, as the figures show, except for the slurry-preparation phase. Lime is transported from a storage area to a tank, where it is slaked with recycled pond water. The resultant lime slurry is then diluted with additional pond water and pumped to a slurry feed tank. Slurry from this tank is further diluted with pond water before being fed to the scrubber. The remaining portions of the two processes are essentially identical.

By a strong and growing margin, utilities prefer the limestone to the lime process. Reference 1 found that approximately 55% of the calcium-based FGD capacity currently operational utilizes the limestone process. When those units currently under construction and planned are also considered, the corresponding value climbs to 59%. The chief reason for this preference appears to be the greater availability of limestone than of lime. As a result of this greater availability, the overall cost of a limestone-based process is generally below that of a lime-based system. Thus, even though a lime system offers a more reactive alkaline material, with resultant advantages in system design and control and reagent utilization, the lower overall cost of a limestone system gives it the preferred status indicated above.

A flow diagram of the materials in the scrubber section of a limestone FGD system is shown in Fig. 2.3. Figure 2.4 presents the same information for a lime-based system, with the calciner (or slaker) also included. These diagrams (which are taken from Ref. 2) are based on an 800-MWe power plant burning high-sulfur (3.5%) Illinois coal. It may be noted that the limestone balances are based on an SO_2 removal efficiency of 90% whereas a value of 95% is used for the lime process. These figures represent current limits of the state-of-the-art technology.

2.1.2 Process Chemistry

Although considerable experience has been gained in the operation of lime/limestone FGD systems, the chemistry involved in these systems is not completely understood. For the limestone process, however, the basic reaction is the combination of the calcium ions in the slurry with the sulfite ions, i.e.

$$Ca^{++} + SO_3^{=} + 1/2\ H_2O \rightarrow CaSO_3 \cdot 1/2\ H_2O \qquad (2.1)$$

The calcium sulfite hemihydrate produced in this reaction is insoluble in water.

Some oxidation of sulfite to sulfate does occur so that gypsum is formed via the reaction

$$Ca^{++} + SO_4^{=} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \qquad (2.2)$$

The chemistry of the lime process is essentially the same with the exception that the slaking reaction is also included in the overall process. This reaction is simply

$$CaO + H_2O \rightarrow Ca(OH) \qquad (2.3)$$

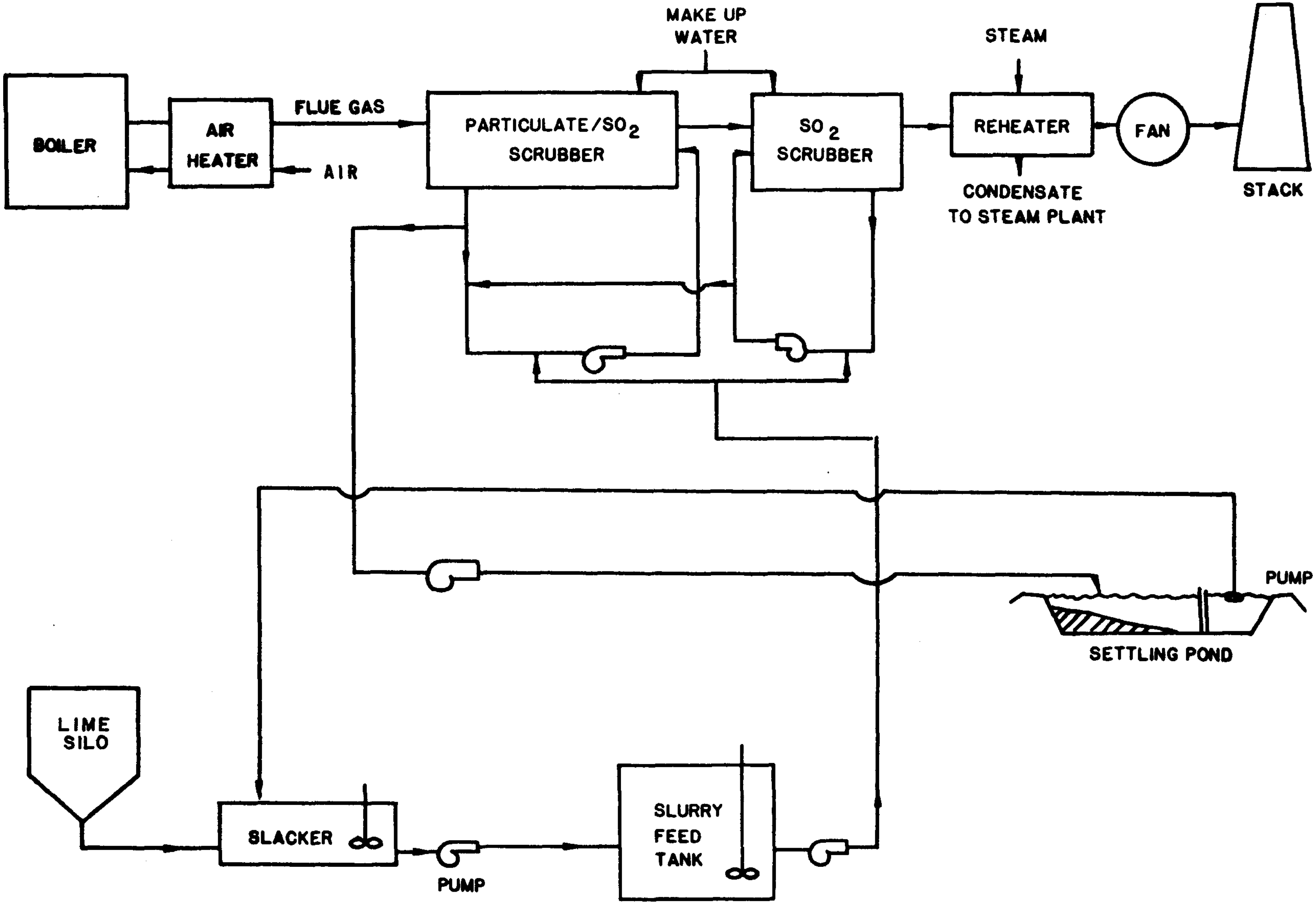


Fig. 2.2 Flow Diagram of Lime Wet-Scrubbing FGD Process

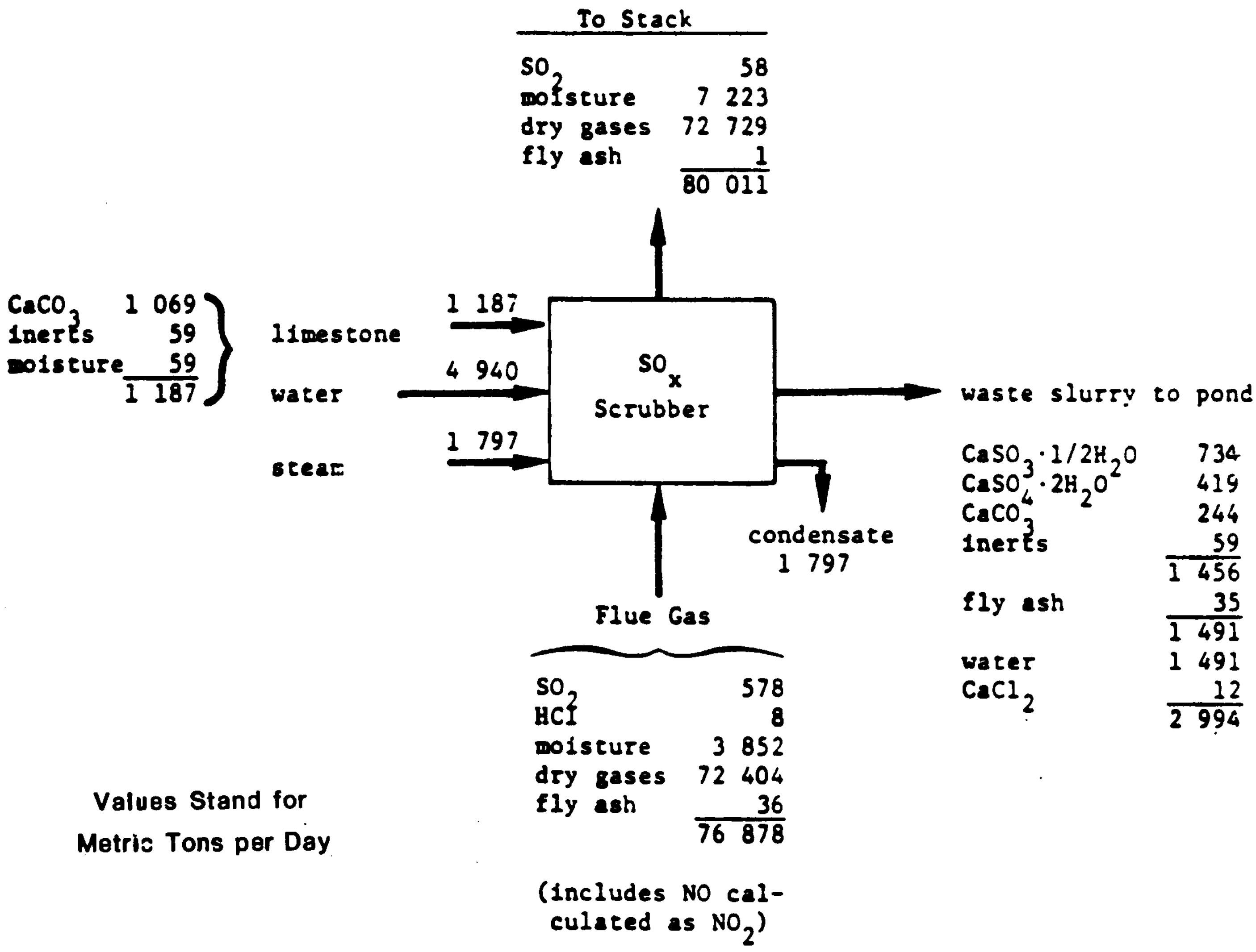


Fig. 2.3 Material Balance of a Limestone Scrubber (Based on 90% Removal of SO_x, 77% Utilization of Ca, and Oxidation to 30% Sulfate)(Source: Ref. 2)

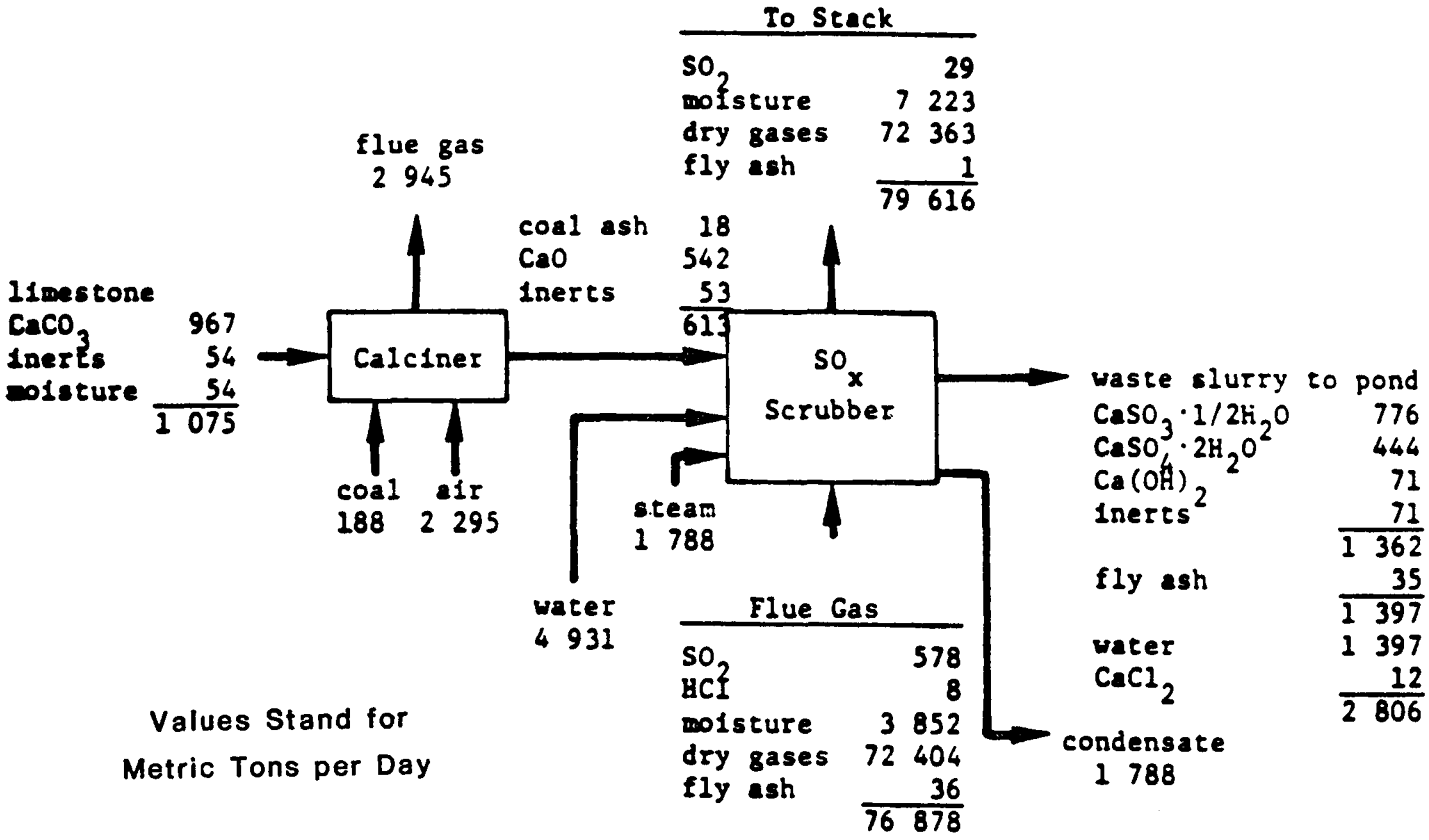


Fig. 2.4 Material Balance of a Lime Scrubber (Based on 95% Removal of SO_x, 90% Utilization of Ca, and Oxidation to 30% Sulfate) (Source: Ref. 2)

The additional hydroxide ion from the slaked lime makes the pH of lime slurries higher than that of limestone slurries. Because of this higher pH, precipitation of $CaCO_3$ is possible via the reactions

$$CO_2 + H_2O \rightarrow 2H+ + CO_3 \qquad (2.4)$$

$$Ca(OH)_2 + CO_3^{=} \rightarrow CaCO_3 + 2OH^- \qquad (2.5)$$

2.1.3 Process Performance

SO_2 Removal Efficiency. As noted above, current technology puts a limit of about 90% removal for the limestone process and about 95% for the lime process. A long-term average for either process may however be only about 85% or less.[3] For purposes of comparison with advanced FGD systems described in subsequent sections of this report, however, the conservative values given in Ref. 4 will be used as the base-case SO_2 removal efficiencies. For the limestone system, therefore, the SO_2 removal efficiency is expected to reach about 80% and for the lime system it can be greater than 90%.

System Reliability. These two processes have had several years of combined operating experience. During this time, many operating difficulties have been identified and various solutions to them proposed. Although several significant problem areas such as corrosion and scaling remain, the accumulation of operating experience and continuing research and development efforts should raise the reliability of lime/limestone FGD systems.

Reference 4 notes that system availability (defined as the ratio of the number of hours the system was available for operation to the total number of hours in the period) can be as high as 90-95%. However, such a high availability requires heavy manpower commitments for cleaning and maintenance.

Sulfur Recovery. Both the basic limestone and lime scrubbing processes are considered to be "throwaway" systems, that is, the sulfur removed from the flue gas cannot be recovered.

Effect of Coal Type on Performance. As noted in Ref. 1, mass-transfer limitations tend to limit the applicability of the most preferred concept (spray towers, as opposed to the venturi scrubbers commonly used) to low- and medium-sulfur coals. When high-sulfur coal is burned, special reagents such as magnesium additives must be used to compensate chemically for the mass-transfer limitations.

That apparent effect of coal type on the performance of a given lime/limestone system has a further effect, therefore, on the system's initial capital investment and subsequent operating costs. For example, coal with a higher sulfur content will require more sorbent and will aggravate corrosion and scaling problems. System reliability could also become even poorer than it is now.

Effects of Plant Size and Operating Mode on Performance. The basic lime/limestone technology and performance is unaffected by plant size or operating mode. These systems can be used in either new or retrofit applications.

2.1.4 Water, Chemical, and Power Requirements

The required inputs for lime/limestone scrubbers obviously depend upon such factors as specific design characteristics of a given unit, sulfur content of coal, SO_2 removal requirements, waste-disposal technique, and many others. For the specific material balances of a limestone and a lime scrubber (shown, respectively, in Figs. 2.3 and 2.4), the scrubber is designed for an 800-MW power plant burning Illinois coal containing 3.5% sulfur.

A typical ratio of liquid to gas in a scrubber is in the range of 60-100 gal/10^3 actual cubic feet (acf), with the higher number being more typical for high-sulfur coals. Most of the liquid, however, is recycled; and the actual consumption of water is typically 1.1 to 1.4 gal/min per MW for high-sulfur coals. This ratio may be lowered, somewhat, to about 1 gal/min per MW for low-sulfur coals. Limestone requirements for a wet-scrubber FGD system are typically of the order of 0.05 to 0.07 ton/MWh for high-sulfur coals. This ratio may be reduced to approximately 0.01 ton/MWh for low-sulfur coals.

The energy requirements for a limestone FGD system consist of two major components: the electrical energy to operate pumps, fans, etc.; and the energy used to reheat the flue gas as it exits the scrubber and prior to discharge from the stack. Electrical requirements are in the range of 1-2% of net electrical output, or 0.4-0.8% of heat input. Flue-gas reheat requirements typically constitute about 2% of the boiler input heat. Thus the total energy associated with a limestone FGD system typically falls in the range of 2.5-3% of boiler heat input. For a lime-based system the energy requirements are increased to 4-5% -- lime preparation requires the additional power. These energy estimates do not include the energy associated with mining, or transportating the raw material, or disposing of the waste from the scrubber.

2.2 ENVIRONMENTAL IMPLICATIONS

2.2.1 Effluents

The effluents released from the stack of a plant equipped with a limestone or lime scrubber are included, respectively, in Figs. 2.3 and 2.4. These figures also give the composition of the waste sludge material produced in each scrubber. For each process the sludge is approximately 50% water. The amount of wet sludge generated in a lime or limestone FGD system with high-sulfur coal is typically on the order of 0.2 ton/MWh.

2.2.2 Waste Disposal

Disposal of the waste sludge from lime or limestone scrubbers is considered by many people to be one of the dominant drawbacks of this technology. These sludges tend to be physically unstable and difficult to dewater. It is possible in some cases to modify the process (at additional expense) to produce gypsum, which can be more easily dewatered. Fixation by adding fly ash and lime, to yield a material like low-grade cement, is another option. An oxidation step between the scrubber and the thickner can also be used to convert the sulfite to sulfate, thereby preventing the solid waste from becoming thixotropic.

Land requirements for sludge ponding (or gypsum storage) also constitute a major problem. In many cases, the ponds must be lined with an impervious clay-like material to prevent toxic chemicals from leaching from the pond into groundwater.

An extensive study of FGD waste-disposal techniques is discussed in Ref. 5. Several generic categories of sludge disposal are examined. These are: untreated ponding, fixation by addition of purchased chemicals; blending of sludge and fly ash without additives; the production of gypsum; and mine disposal of a sludge-fly ash mixture. Most of the fixation processes lead to ultimate disposal in landfills. An economic assessment was made for each of the systems as applied to a 500-MW plant burning Illinois coal of 3.5% sulfur and 16% ash. The scrubber was assumed to operate with a stoichiometry of 1.5 and produce a waste product composed of 15% solids.

Capital-cost estimates associated with waste disposal ranged from $15-48/kW (mid 1979 dollars). The low estimate was associated with the gypsum-production process, and the high estimate was for one of the chemical-fixation processes. Other fixation techniques would require capital investments of $21-27/kW. The conventional process of ponding the untreated sludge in an earthen-diked pond was estimated to have a capital-cost requirement of $34/kW. First-year revenue requirements were lowest for the sludge-ponding technique, with a value of 0.9 mill/kWh. The gypsum-production process was estimated at 1.2 mill/kWh, and the other processes ranged between 1.5 and 2 mill/kWh. These estimates were made under the assumption of an 80% capacity factor. When expressed in terms of $/ton of dry waste, the above first-year revenue requirements become: $8.1 for the sludge-ponding technique, $7.9 for the gypsum-production process, and $8.2 for the mine-disposal process. The sludge-fly ash blending process would cost $9.3/ton of dry waste, and the chemical-additive processes would range from $12-16/ton.

Reference 5 also examined the effects of plant size, sulfur content, ash content, and distance to disposal site for these various disposal processes. The relationship to plant size is the typical one, with the capital cost leveling out at approximately 800 MW. For the conventional sludge-ponding technique, the estimated capital cost for an 800-MW plant is approximately $28/kW as compared to $34/kW for a 500-MW plant. The first-year revenue requirements show a similar relationship to plant size, with the sludge-ponding processes leveling out at approximately 0.9 mill/kWh for plants having a capacity of 500 MW or more. The gypsum-production and chemical-additive processes show a more rapid decrease in revenue requirements as plant size increases -- they approach the sludge-ponding costs for plants larger than about 1000 MW. The sulfur and ash contents affect the water disposal costs in a linear fashion, although the rates of increase (i.e., the slopes of the curves) are different for each process. An increase in sulfur content to 5% (from the assumed 3.5% level) would increase the capital costs of the sludge-disposal process to $40/kW, and the first-year revenue requirement to 1.3 mill/kWh. Increasing the ash content from 16 to 20% would also increase the capital costs to about $40/kW when the conventional sludge-ponding process is used -- and this would slightly increase revenue requirements, to an expected 1 mill/kWh.

The distance between the plant and the disposal site has a very significant impact on the costs when a conventional sludge-ponding process is used. Increasing this distance from the 1 mi assumed in the base-case analysis to 10 mi would increase the capital investment to about $73/kW -- more than double the base-case value. First-year revenue requirements would also more than double, to about 2 mill/kWh. On the other hand, the landfill processes show only a weak dependence of capital cost on distance, although revenue requirements obviously increase because of increased transportation costs.

The information summarized above shows that the economics of waste disposal can have a pronounced effect on the overall capital investment and operating costs of a lime/limestone FGD system. Differences in disposal techniques and site-specific parameters (e.g., distance to disposal site, coal composition, etc.) can lead to dramatically different costs. These effects will be discussed further in the next section of this report.

2.3 ECONOMIC AND MARKET FACTORS

2.3.1 Operating and Capital Costs

The wet-scrubber FGD technology using lime or limestone sorbent was selected as the reference technology for utility application because it is the most developed technology in terms of installed capacity. As noted earlier, the dominance of this technology is expected to continue into the near-term future. However, the lime/limestone FGD technology cannot truly be considered as a "mature" technology. Changes in the process technology which have prevented it from becoming truly mature have resulted from several factors. One of the contributing factors to such changes is the operating experience with the technology. As experience is gained with wet-scrubbing systems, problems are identified and design or operational corrections are proposed. For example, plugging problems have been instrumental in the evolution of the scrubber design. Another factor which can necessitate changes in the technology is the waste-disposal problem discussed in Sec. 2.2.2. The disposal technique affects the basic technology by requiring design and/or operational changes. For example, the production of gypsum requires that air-oxidation equipment be added to the scrubber loop. The various fixation or chemical-additive waste-disposal processes require other design and operational changes. A third factor forcing alterations in the technology consists of the changing federal and state requirements for SO_2 and particulate removal. As these requirements become more strict, various changes in the technology must be made to comply with them. Other factors which can affect the maturity of the technology include the variation in site specific conditions such as coal type, sulfur content, ash content, sorbent and water availability, etc. Because such conditions vary greatly, the experience gained at one site is sometimes difficult to translate to another site.

Each of the above factors (as well as others not noted) has had an effect on the evolution of the lime/limestone FGD technology. Naturally, the capital and operating costs have also been affected. Costs are also affected by nontechnical factors. These include availability and productivity of labor, interest rates, distance to disposal site, allowance for funds during construction, and availability of sorbent.

The cost variations for many existing lime/limestone FGD systems have been examined in some detail and are reported in Ref. 6. In this reference, attempts were made to normalize those cost estimates to a consistent set of conditions and to develop relationships of cost to various factors such as plant size and sulfur content. Although this work goes a long way in providing more consistent cost data, we have elected not to base our cost estimates on those plants already operating or under construction but rather to base them upon more current estimates. The more up-to-date estimates, we believe, will more accurately reflect capital and operating expenditures associated with future lime/limestone FGD systems.

Reference 7 presents data for a new two-unit coal-fired plant producing 1000 MW (net). This facility was assumed to be located at a midwestern site. Both lime and limestone FGD sysems were evaluated. Estimates were made for a high-sulfur (4%) eastern coal and a low-sulfur (0.48%) western coal. In all cases, the waste was thickened with fly ash and additional dry lime to produce a mixture composed of at least 70% solids by weight. This waste was then trucked to a landfill area 10 mi from the power plant. To meet the emission standards assumed in this study, the SO_2 removal efficiency was designed to be 87% for the high-sulfur eastern coal and 85% for the low-sulfur western coal.

The cost estimates, escalated to beginning-of-year (BOY) 1980 dollars, are shown in Table 2.1 for each case. These estimates include waste disposal but do not include particulate removal. As shown in this table the capital requirement for a limestone system is about 6% greater than that for a lime system. This difference is attributed to the lower material-handling rates resulting from the lower molecular weight, higher reactivity, and lower stoichiometric ratio of the lime system relative to the limestone system. The slightly greater variable cost of operation and maintenance (O & M) for the lime system reflects the higher cost of the sorbent raw material. The capital costs for a low-sulfur coal application are approximately 75% of the corresponding value for a high-sulfur coal system for both lime and limestone systems. Variable O & M costs for the low-sulfur case are less than half of those for the high-sulfur cases, thus reflecting the lower sorbent requirements.

Estimates are presented in Ref. 8 for a 500-MW plant burning eastern coal with a 3.5% sulfur content and equipped with a limestone FGD system. These results have been escalated to BOY 1980 dollars and are presented in Table 2.2. The SO_2 removal efficiency was defined to be "at least 90%" in this study. The variable O & M cost includes waste disposal at $11/ton.

Table 2.1 Cost Estimates for Two 500-MW Wet-Scrubbing Systems, One Using Lime and the Other Using Limestone (BOY[a] 1980 $)

	Size of Expenditure			
	Lime System[b]		Limestone System[b]	
Required Expenditure	High-Sulfur Coal	Low-Sulfur Coal	High-Sulfur Coal	Low-Sulfur Coal
Capital investment ($/kW)	93	71	98	75
Plant investment ($/kW) (includes general facilities, fees, and contingencies)	132	101	140	107
Total capital requirement ($/kW) (includes preproduction costs and allowance for funds during construction)	168	125	178	133
First-year fixed O & M ($/kW-yr)	9.4	7	10	7.5
First-year variable O & M (mill/kWh)	3.7	1.5	3.2	1.5

[a] Beginning of year.

[b] Includes a spare FGD module.

Adapted from Ref. 7.

Table 2.2 Cost Estimates for a Single 500-MW Limestone System[a] (BOY 1980 $)

Required Expenditure	Size
Capital investment ($/kW)	72
Plant investment ($/kW)	101
Total capital requirement ($/kW)	122
First-year fixed O & M ($/kW-yr)	7.1
First-year variable O & M (mill/kWh)	3.8

[a] Does not include a spare FGD module.

Adapted from Ref. 8.

A comparison of the data in Table 2.2 with those in Table 2.1 shows some unexpected results. The capital costs and fixed O & M costs for the single 500-MW plant are much lower than for the two 500-MW units at the same site. The difference in the capital cost estimates is primarily due to different design concepts. As a means of increasing system reliability, the current design philosophy is to include an extra FGD module in the system. Typically, this means having four modules, each designed for 33-1/3% of the plant output. The estimates in Table 2.1 were based on this current philosophy, whereas the data in Table 2.2 were not. Based on a relationship developed in Ref. 6, it is estimated that the capital cost of a 500-MW limestone FGD system would increase 28% with the inclusion of a fourth, redundant module. An increase of this magnitude would make the capital-cost data in Table 2.2 more consistent with those in Table 2.1. For example, the total capital investment of $122/kW would be increased to $156/kW, which compares more closely to the $178/kW reported in Table 2.1.

Reference 5 also includes estimates for a 500-MW plant burning 3.5%-sulfur coal. These estimates are based on the most recent design concepts. The total capital requirement for the limestone FGD system is estimated to be $158/kW (BOY 1980 dollars). The fixed O & M costs for this plant are estimated at $13.8/kW-yr and the variable O & M costs at 1.4 mill/kWh. The fixed costs are somewhat higher and the variable costs somewhat lower than the previous estimates. It is not clear why these differences arise, but it may be attributable to slightly different splits among the O & M costs.

Another set of cost estimates are presented in Ref. 2. Estimated capital costs of $140 kW and $147/kW, for the lime and limestone processes respectively, are reported (with values escalated to BOY 1980 dollars). These estimates are representative of an 800-MW plant burning eastern coal of 3.5% sulfur. The fixed and variable O & M costs for the lime system are $7.1/kW-yr and 2 mill/kWh. Corresponding values for the limestone sytem are $7.4/kW-yr and 1.5 mill/kWh. The variable O & M costs reported in this reference again seem to be somewhat low, but insufficient data were available to determine the cause.

The data presented above show the capital cost for a modern lime FGD system, in a large plant burning high-sulfur eastern coal, ranging from $140-168/kW. A range of $147-178/kW was reported for a comparable limestone system. Fixed O & M costs ranged from $7.1-9.4/kW-yr for the lime system and $7.1-13.8/kW-yr for the limestone system. Variable O & M costs ranged from 2-3.7 mill/kWh and 1.5-3.8 mill/kWh, respectively. A plant having a 70% annual capacity factor would be expected to have total O & M costs ranging from 3.2-5.2 mill/kWh for a lime system and 2.7-5 mill/kWh for a limestone system. Low-sulfur applications would have capital costs equal to about 75% of the high-sulfur design and variable O & M costs of less than half those of the high-sulfur case.

The data and analyses presented in this section indicate the costs associated with flue-gas desulfurization are inherently variable. Several complicating factors have been discussed: design philosophy, waste-disposal techniques, sulfur content, SO_2 removal requirements, and others. Many of these factors will also appear when estimating the costs of more advanced FGD systems. Therefore, wherever possible, we will present cost estimates of the advanced technologies side by side with estimates for the reference technology so that the relative costs can be observed even in those cases where the "real costs" of the advanced system cannot be reported.

2.3.2 Potential Applications

The basic lime/limestone technology can be applied to all types of coal and to either utility or industrial units. As noted earlier, high-sulfur coal requires a slightly different scrubber design than does low- or medium-sulfur coal, but the basic technology is the same for all three coals.

2.3.3 Market Potential

The market potential of this base-case FGD technology is reflected in the installed capacity to date and in the capacity which has been contracted for but not yet installed. These figures were presented in Sec. 2.1.1.

2.4 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT

Numerous research and development programs are underway to improve lime/limestone FGD systems with respect to costs, reliability, and waste handling. One such effort is discussed in Ref. 9, which reports that the use of fly ash as the alkaline source makes it possible to significantly reduce the amount of lime or limestone added to the slurry. A pilot system operated by Peabody Process Systems for the Minnesota Power and Light Co. allowed SO_2 emission standards to be met while using only the alkali in the fly ash, i.e., with no supplemental lime or limestone. Economic benefits for the use of fly ash alone were indicated by this pilot program to include the following.

- Reduction (or under optimum circumstances, elimination) of alkali cost.
- Reduction in power consumption for the FGD system.
- Reduction in initial capital investment.
- Reduction in solid-waste handling and disposal cost.
- Improved system reliability, principally by elimination of the scaling problem common in conventional lime/limestone systems.

Other methods of improving the basic lime/limestone FGD system include the injection of magnesium or adipic acid into the scrubber loop. Each of these additives acts to accelerate the mass transfer at the gas/liquid interface, thereby enhancing the SO_2 removal efficiency. Additional research and development efforts are directed towards sludge-disposal techniques and improvements in the mechanical design of various components, e.g., mist eliminator, reheaters, or basic scrubbers. These actual or potential improvements in the basic lime/limestone technology will not be further elaborated upon in this report. They have been noted only to reflect some of the research and development efforts currently underway in the FGD base technology system.

2.5 REFERENCE FGD TECHNOLOGY FOR INDUSTRIAL-BOILER APPLICATIONS

Although this report is primarily concerned with utility FGD applications, SO_2 must be removed from flue gas produced in industrial boilers as well. A federal standard of 1.5 lb of SO_2 per 10^6 Btu has been proposed for industrial boilers.[10]

In contrast to utility uses, where the vast majority of the FGD systems currently use either the wet-lime or wet-limestone throwaway process, small industrial boiler applications use soluble sodium-based systems almost exclusively. This choice is predicated by the highly reliable scrubbers and relative operational ease associated with soluble systems. The availability of process streams as suitable scrubbing liquor also favors a once-through process of this type. This process is also responsive to changes in boiler-load conditions.

The basic process employs a solution of NaOH, Na_2CO_3, or $NaHCO_3$ to absorb SO_2 from the flue gas and to produce Na_2SO_3, Na_2SO_4, and $NaHSO_3$. Removal efficiencies of 90% are achieved with this process. The scrubber effluent can then be treated in a number of ways. Some fresh caustic or solid ash can be added to the effluent, which is recycled with a slip stream going to a wastewater treatment facility and ultimate disposal. Some industries have process streams available that can be used as the liquor source. Effluent from these processes is typically not recycled. Due to the high solubility of the sodium-based salts, the waste-disposal techniques associated with this process are of great concern and are rather costly.

Reference 11 presents capital and operating cost estimates for various industrial FGD processes. For a $100 \cdot 10^6$ Btu/h boiler burning coal with a 3.5% sulfur content and with an SO_2 removal requirement of 90%, the estimated capital cost (in 1980 dollars) is reported as $1,060,000. A boiler four times as large would double that capital cost, to approximately $2,120,000. Systems of similar size, burning 0.6%-sulfur coal and expected to remove 75% of the SO_2, would have capital costs of approximately $680,000 ($100 \cdot 10^6$ Btu/h) and $1,130,000 ($400 \cdot 10^6$ Btu/h). Annual O & M costs for the 3.5%-sulfur cases were estimated at $440,000 for the 100-million rating and $1,390,000 for the 400-million rating. The low-sulfur applications (0.6%) have estimated annual O & M costs of $315,000 for the 100-million rating and $570,000 for the 400-million rating. The energy requirement associated with a sodium-scrubber industrial FGD system is approximately 2-2.6% of the net heat input to the boiler.

Where available, data on industrial applications for the advanced FGD technologies considered in this report will be presented. The sodium-based throwaway process will serve as the reference technology because of its predominant utilization in the industrial sector.

3 DRY INJECTION/BAGHOUSE FILTERING

SUMMARY

The dry injection/baghouse filtering technology is a comparatively simple mechanical system with a comparatively low capital investment. Low operating and maintenance costs and a low rate of forced outages can also be expected from this technology. Because it is a dry FGD process, the water requirements are very much less than the base case lime/limestone technologies. One disadvantage of this system is a somewhat low SO_2 removal efficiency, which may limit its application to low-sulfur coals. Additional potential problems include dependence on a limited supply of nahcolite and the need to dispose of a relatively soluble sodium-based salt, which could leak into underground water supplies. This FGD technology is in the testing stage, with prototypes at 60-MW and 80-MW stations.

3.1 TECHNICAL ASPECTS

3.1.1 Process Description

The simplest dry FGD system in terms of number and complexity of components is a powder/bag filter system in which the powder (e.g., nahcolite) is allowed to come in contact with the flue gas. This produces a dry salt as the final disposable product.

A simplified flow diagram for such a system, with nahcolite used as the sorbent material, is shown in Fig. 3.1. Nahcolite ore is a naturally occurring material that contains from 70% to 90% sodium bicarbonate. After arriving at the plant site, the nacholite ore is crushed to 1/4-in rock. The crushed rock is then fed to a dry ball mill, where it is pulverized (90% through 200 mesh). Some of the ground particles (typically 20%) are fed into a stream of flowing air and carried to precoat the filter bags. This precoat technique is used to increase the effective residence of the sorbent in the flue-gas stream and thereby increase the SO_2-to-sorbent utilization ratio. The remaining portion of the ground nahcolite ore is slug-fed as rapidly as possible into the flue-gas stream just upstream of the baghouse. An alternative to the slug-feed method is to continuously feed the ore into the stream at a controlled rate.

A baghouse usually has several compartments, each of which contains many individual bags, or filters. A typical bag has a 1-ft-diameter, 30-ft-high cylinder open at each end. It is constructed of a heat-resistant synthetic fiber such as fiber glass.

The flue gas is drawn into the baghouse by induced-draft fans. At this point, the gas contains the finely ground nahcolite ore and some fly ash. Some, but not all, systems incorporate an electrostatic precipitator just

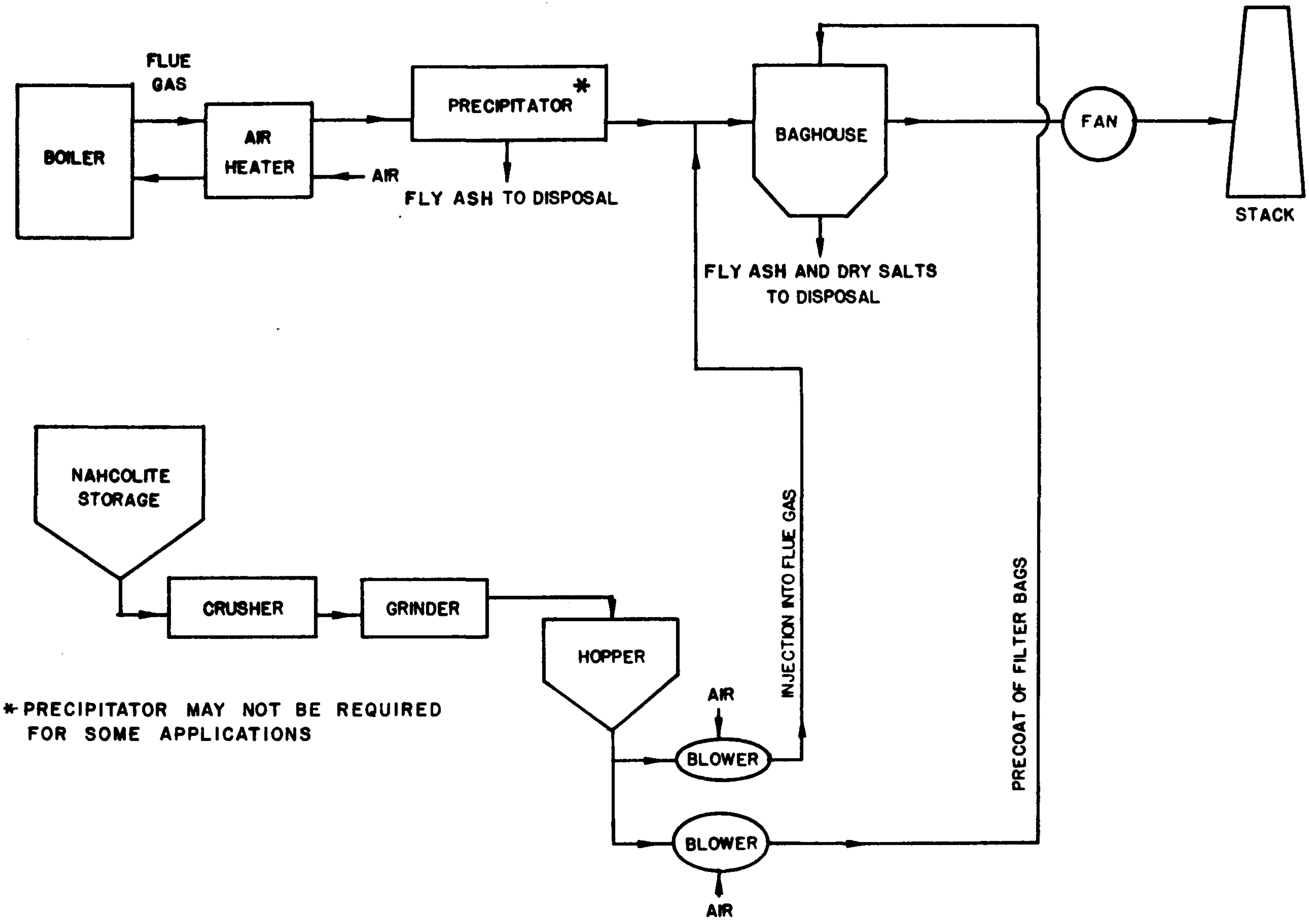


Fig. 3.1 Flow Diagram of Dry Injection/Baghouse Filtering FGD Process

upstream from the point of nahcolite injection. However, at least some fly ash would enter the baghouse along with the flue gas and the nahcolite ore.

As the flue gas enters a compartment of the baghouse it flows through the bag wall into an external plenum chamber. The nahcolite and fly ash form a layer of deposits that remain on the inside bag surface. This layer of solid, including the nahcolite which has been precoated on the bags, becomes the site in which the SO_2 is reacted and upon which the fly ash is deposited.

After a specified time period, the filters are cleaned via reverse air, a high-pressure air pulse, or a mechanical shaking procedure -- and the solid sodium salts and fly ash are removed for disposal. Depending upon local conditions, the waste is taken directly to a landfill site or processed for insolubilization prior to landfilling.

3.1.1.1 Test Programs

Several test efforts have been undertaken over the years to evaluate the baghouse-filtering process as a viable flue-gas-desulfurization technique. Some of these efforts are summarized below. This information was obtained principally from Refs. 12-16.

A parameter of significance in the use of the dry-sorbent baghouse process is the stoichiometric ratio. This parameter is defined as the ratio of the sodium bicarbonate in the nahcolite fed into the system to the amount theoretically required to react with all the sulfur dioxide in the flue gas. Equivalently, the stoichiometric ratio can be expressed as the ratio of the mole equivalent of the Na_2O in the nahcolite ore to the moles of SO_2 in the flue gas. A typical relationship between stoichiometric ratio and percent SO_2 removal is shown in Fig. 3.2. (This figure was taken from Ref. 15.) From this figure, it is apparent that to obtain SO_2 removal efficiencies greater than about 70% -- when the flue gas is approximately 275°F as it exits a typical air preheater on an existing coal-fired power plant -- the stoichiometric ratio must be 1.5 or greater. With such a high stoichiometric ratio, the nahcolite-utilization factor will decrease -- and that decrease will increase the quantities of ore required and thus the annual operating costs of the FGD system. It is also apparent from Fig. 3.2 that an increase in flue-gas temperature can be advantageous to SO_2 removal: a higher temperature can reduce the necessary stoichiometric ratio for a given SO_2 removal efficiency. This feature is shown in Fig. 3.3 (again from Ref. 15), where the curve indicates that high SO_2 removal efficiencies can be attained with a reasonable stoichiometric ratio if the gas temperature can be increased. For example, it appears that 90% removal can be achieved with a gas temperature of about 500°F.

Another significant parameter in this FGD technology is the air-to-cloth ratio, or superficial velocity. This parameter is defined as the flue-gas flow rate divided by the bag area. It is expressed in terms of cubic feet per minute for each square foot (cfm/ft^2) or equivalent units. Test results indicate that SO_2 removal will decrease as the air-to-cloth ratio increases. This conclusion is not unexpected because the air-to-cloth ratio is indicative of the effective residence time for which the flue gas is exposed to the filters.

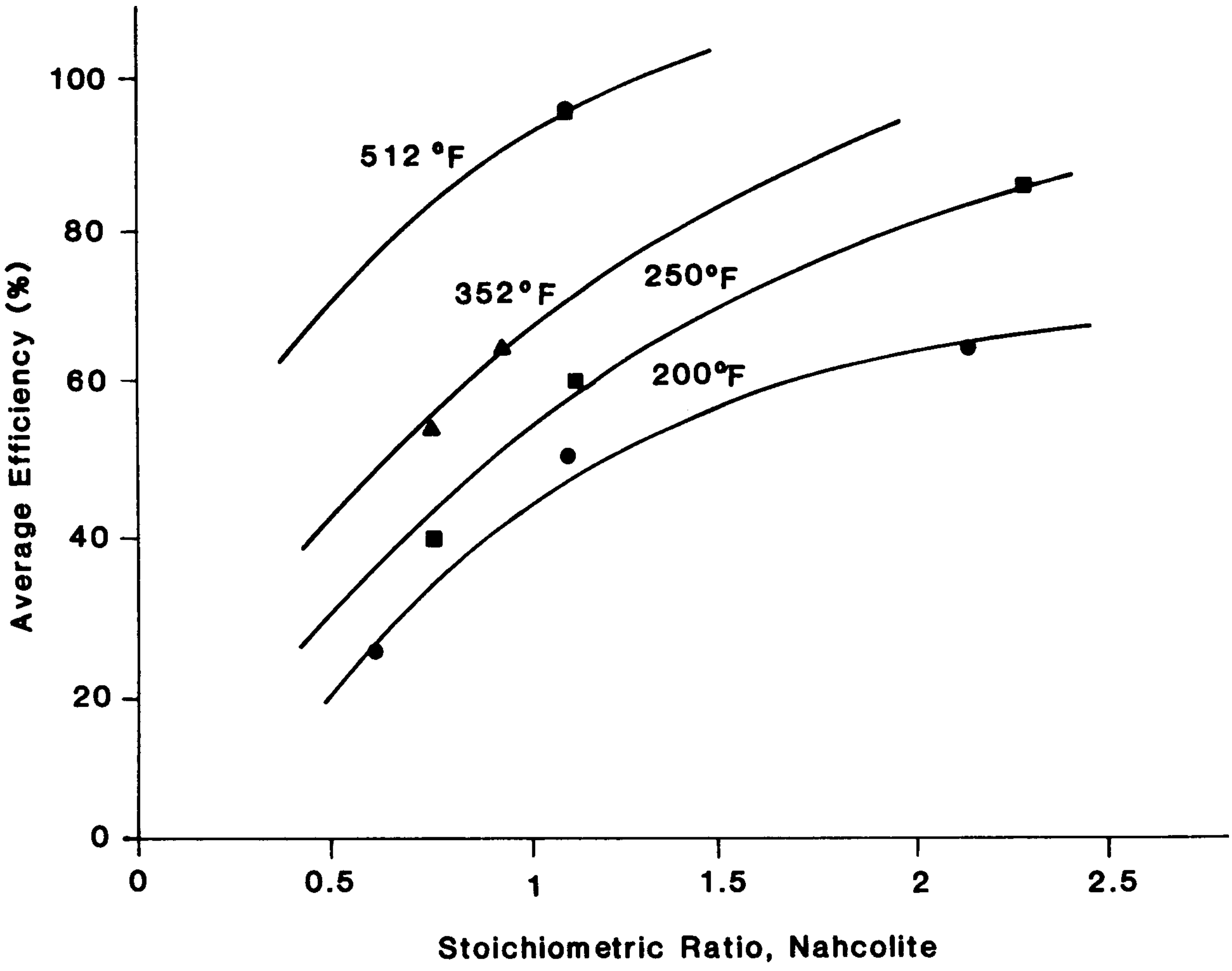


Fig. 3.2 Combined Effect of Stoichiometric Ratio and Flue-Gas Temperature on SO_x Removal Efficiency in Dry Injection/Baghouse Filtering (Source: Ref. 15)

In 1965 a baghouse FGD system was tested at the 320-MW Los Alamitos station, belonging to Southern California Edison. At that time, the facility burned 1.5%-sulfur oil. For this test, nahcolite ore was injected into the flue gas at stoichiometric ratios between 0.5 and 1. The gas flow rate ranged from 700 to 1500 actual cubic feet per minute (acfm) and was at temperatures around 240°F. SO_2 removal efficiencies of 48% to 73% were achieved with air-to-cloth ratios of 3-6.5 ft/min. As expected, the higher removal efficiencies were obtained with the lower air-to-cloth ratios. The use of nahcolite was not further pursued in this test because of a lack of availability. The difficulty in obtaining significant quantities of nahcolite ore will be a recurring theme in the discussion of baghouse-filter FGD systems. Successful SO_2 removal was also achieved in this program using dolomitic limestone as the sorbent material.

Pilot-scale tests were conducted in 1967-1969 by Wheelabrator-Frye at the Edwardsport station of Public Service of Indiana. Some 16 different sorbents were examined, but only Na_2CO_3 (soda ash) and $NaHCO_3$ were found to be

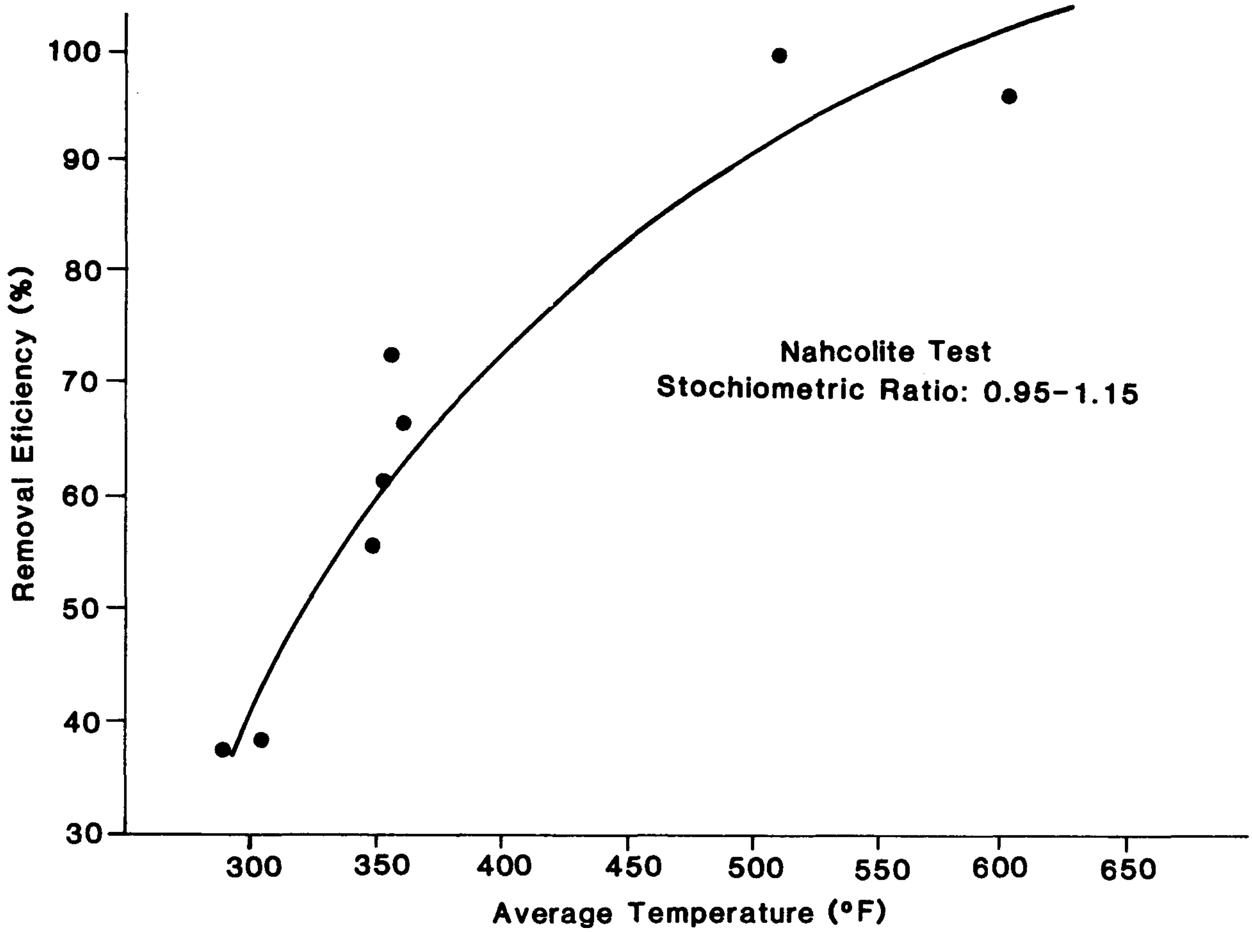


Fig. 3.3 Effect of Temperature on SO_2 Removal Efficiency in Dry Injection/Baghouse Filtering (Source: Ref. 15)

consistently effective for SO_2 removal. Experimental SO_2 removal ranged from 13% to 72%; but the higher removal was achievable only at unacceptably low sorbent-utilization factors, which ranged from 22% to 93%. This test effort was of such small scale, however, that the operating experience obtained from it is not considered to be applicable for current utility situations.

Owens-Corning Fiberglass Corp. investigated SO_2 removal in an eight-bag baghouse at gas flow rates between 300 and 400 acfm. High air-to-cloth ratios of 6-8 ft/min and low stoichiometric ratios of 0.7-0.8 were used in these tests, which were conducted at 300°F. The reported SO_2 removal efficiencies obtained in these tests were about 60%.

The Air Preheater Co. tested several sorbents -- including commercial sodium bicarbonate, nahcolite, and hydrated limes -- at the Public Service Electric and Gas Co. Mercer station in New Jersey. These tests, which were conducted during 1968-1969, examined such variables as the mode of baghouse operation (cyclic or parallel) and the air-to-cloth ratio. Gas flow rates were in the range of 7,500-15,000 cfm. Most of the tests were run at gas-inlet temperatures between 300°F and 350°F, although some tests between 500°F and 600°F were also conducted. It was concluded from these tests that lime is not an adequate dry sorbent for SO_2 removal, but that nahcolite and commercial $NaHCO_3$ performed well. The data from these tests showed that removal efficiencies of 60-80% could be obtained with gas temperatures of 300-350°F and stoichiometric ratios between 0.7 and 1.8. To obtain maximum SO_2 removal, however, gas temperatures of greater than 500°F would be needed.

In 1974, Superior Oil tested a bench-scale fired bed of nahcolite at the Cherokee station of Public Gas Co. These tests showed 80-95% SO_2 removal and a sorbent utilization of greater than 90%.

In 1976, Bechtel Corp., under contract to the Electric Power Research Institute, conducted a survey of dry-sorbent usage for flue-gas desulfurization. The results of this survey are in agreement with many of the conclusions expressed in the tests mentioned above. The principal conclusions are that lime is relatively ineffective for baghouse FGD application and that, of several other sorbents, only $NaHCO_3$ and Na_2CO_3 were effective enough to warrant further consideration as dry sorbents for FGD applications. It was also concluded that, in general, the higher the flue-gas temperature, the greater the SO_2 removal. The overall conclusion of this survey was that injecting baghouses with nahcolite appeared to be the most promising method of FGD using a dry sorbent.

3.1.1.2 System Flow Diagrams

Reference 14 provides conceptual data for two steam electric plants equipped with nahcolite/baghouse filter FGD systems. One plant is assumed to be a coal-fired boiler (800 MW) in the southwestern United States burning pulverized low-sulfur (0.7%) western coal. The other plant has two 840-MW cyclone boilers located in the midwestern United States and burning typical midwestern high-sulfur (3.5%) coal. The basic baghouse design for each system is similar. Multiple compartments exist in each baghouse. The southwestern plant requires 10 compartments, but the midwestern plant needs 14 compartments. Each of these compartments contains 14 cells, or hoppers, with 90 bags in each hopper. The bags are 30 ft long, 1 ft in diameter, and are made of silicone-treated fiberglass. The compartments are designed so that one of the 14 cells is being cleaned and another is being maintained at all times. The life expectancy of the bags is assumed to be two years. Some of the parameters for the two systems are listed in Table 3.1. The data in this table represent what are believed to be reasonable estimates and extrapolations of existing information. In formulating the data in Table 3.1, it was assumed that SO_2 removal efficiencies of 70% and 75% were necessary for the southwestern and midwestern plants, respectively.

3.1.2 Process Chemistry

The chemical reactions which occur in the filters are believed to proceed as discussed below.

First the heat from the flue gas decomposes some of the sodium bicarbonate in sodium carbonate via the reaction

$$2NaHCO_3 \rightleftarrows Na_2CO_3 + H_2O + CO_2 \qquad (3.1)$$

Table 3.1 Design Parameters for Dry Injection/Baghouse Filtering Applications in Southwestern and Midwestern U.S. Plants

Parameter	S.W. Plant	M.W. Plant[a]
Boiler capacity (MWe)	800	840
Average load factor (%)	68	65
Coal heating value (Btu/lb)	8,997	10,480
Feed rate (lb/h)	842,000	816,000
Flue-gas temperature at entrance to baghouse (°F)	240	330
SO_2 concentration in gas (ppm)	681	2,893
Gas mass flow rate (lb/h)	$8.04 \cdot 10^6$	$9.07 \cdot 10^6$
Gas flow rate at baghouse (acfm)	$2.35 \cdot 10^6$	$2.99 \cdot 10^6$
Air-to-cloth ratio	2.4	2.2
Air flow to precoat bags (scfm)	$1.79 \cdot 10^5$	$4.25 \cdot 10^5$
Precoat nahcolite rate (lb/h)	12,557	40,030
Air flow for nahcolite injection (scfm)	3,000	9,600
Injection nahcolite rate (lb/h)	50,230	160,100
Injection cycle time (min/h)	15	38

[a] Consists of two boilers, each identically characterized by the values below.

Source: Ref. 14.

The sodium carbonate then reacts with the SO_2 via

$$Na_2CO_3 + SO_2 \rightarrow Na_2\ SO_3 + CO_2 \qquad (3.2)$$

The sodium bicarbonate that has not been decomposed removes sulfur dioxide via the reaction

$$2NaHCO_3 + SO_2 \rightleftarrows Na_2\ SO_3 + 2CO_2 + H_2O \qquad (3.3)$$

Some of the sodium sulfite produced in the above reactions is oxidized to a sulfate, i.e.

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2\ SO_4 \qquad (3.4)$$

3.1.3 Process Performance

Based upon the plant parameters in Table 3.1 and the chemical reactions discussed above, the reaction products given in Table 3.2 can be calculated.

Table 3.2 Chemical Data for Dry Injection/Baghouse Filtering Applications in the Southwestern and Midwestern U.S. Plants (lb/h)

Chemical Datum	S.W. Plant	M.W. Plant[a]
$NaHCO_3$		
Theoretical	31,382	300,128
Stoichiometric ratio · theoretical	47,074	300,128
Nahcolite ore	62,785	400,271
Stoichiometry		
SO_2 removed (Eqs. 3.2, 3.3)	8,378	85,846
Na_2SO_4 formed (Eq. 3.4)	18,581	190,343
$NaHCO_3 \rightarrow Na_2SO_4$ (Eqs. 3.3, 3.4)	21,973	225,153
$NaHCO_3 \rightarrow Na_2CO_3$ (Eq. 3.1)	25,114	75,030
Na_2CO_3 formed (Eq. 3.1)	15,842	47,329
H_2O formed (Eqs. 3.1, 3.3)	5,049	32,184
CO_2 formed (Eqs. 3.1, 3.2, 3.3)	18,086	137,575
O_2 used (Eq. 3.4)	2,093	21,440
Baghouse solids		
Fly ash @ 99.5% removal	4,428	414
Na_2SO_4	18,581	190,343
Na_2CO_3	15,842	47,329
Unreacted ore	15,698	100,088
Total	54,549	338,174
Fly ash and ore escaping (estimate)	53	202

[a] Consists of two boilers, each identically characterized by the lb/h values below.

Source: Ref. 14.

SO_2 Removal. The dependence of the SO_2 removal efficiency on gas temperature and stoichiometric ratio has been previously discussed. Over reasonable ranges of these parameters, SO_2 removal efficiencies in the range of 70-80% appear possible.

System Reliability. No information is available at this time. However, this system is one of the least complex FGD technologies, and it can be anticipated that the reliability would reflect this factor.

Sulfur Recovery. As currently envisioned, sulfur cannot be recovered from this system, which is therefore classified a throwaway cycle.

Effect of Coal Type on Performance. The test program to date has been concerned principally with low-sulfur coals. While the basic technology is applicable to high-sulfur coals as well, an economic penalty must be paid through the use of a higher flue-gas temperature and/or greater sorbent requirements.

Effects of Plant Size and Operating Mode on Performance. No information is available at this time.

3.1.4 Water, Chemical, and Power Requirements

Table 3.2 indicates that the nahcolite ore requirements are 0.04 and 0.12 ton/MWh for the plants in the Southwest and Midwest, respectively. Power requirements can be estimated from information given in Ref. 14. For the unit located in the southwestern U.S., the power requirement for the FGD system is estimated to be about 1% of the electrical power output. For the plant located in the midwestern U.S., the corresponding number is 1.6%. Expressed in other terms the electrical power requirements represent less than 1% of the boiler heat. No reheat of the gas exiting the baghouse is needed. These low power estimates reflect the fact that a dry injection/baghouse filtering system is relatively simple as compared with other FGD technologies.

No specific data are available with regard to the water requirements for this system. The amount of water used however is a small fraction of that needed for a conventional lime/limestone system and would be principally used in the crushing and grinding operations.

3.2 ENVIRONMENTAL IMPLICATIONS

3.2.1 Effluents

The effluents from this FGD process are noted in Table 3.2. The principal environmental concern is the production of sodium-based salts that are relatively soluble and could find their way into the water table. Disposal efforts must be developed with this factor in mind. The amounts of solid waste are shown to be approximately 0.03 and 0.10 ton/MWh for the two cases considered.

Another potential concern is the amount of CO_2 released to the atmosphere from this process. Table 3.2 indicates that approximately 0.01 and 0.04 ton/MWh of CO_2 would be formed in the southwestern and midwestern FGD installations, respectively.

3.2.2 Waste Disposal

3.2.2.1 Landfill

The dry salts produced in this FGD process must be dealt with in an economically and environmentally acceptable manner. Because of the large fixed capital investments and annual operating costs associated with recycling or product processing, the current belief is that the dry salts should be buried in a landfill. Two basic methods of landfill are proposed in Ref. 14. The first method is a combined trench and area landfill procedure. In this procedure, the spent salts are compacted into a cell. This compacted cell is then covered completely with a claylike material, which in turn is also compacted. Each of these cells is placed in a landfill constructed on a base material of low permeability so that ground water cannot leach out the sodium-based salts. About 75,000 tons of baghouse filter cake could be disposed of per acre. Based on the landfill design presented in Ref. 14, it is estimated that for an 800-MWe plant burning 0.7%-sulfur southwestern coal, approximately 2-2.5 acre/yr would be required for disposal of FGD wastes. For the same plant burning 3.5%-sulfur midwestern coal, approximately 10-12.5 acre/yr would be required.

Another method of waste disposal involves insolubilizing sodium sulfate by coprecipitation with acidic ferric ions. The insoluble double salts $NaFe_3(SO_4)_2(OH)_6$ (natrojarosite) and $Na_2Fe(SO_4)_2(OH)\cdot 3H_2O$ (sideronatrite) are produced as the disposable product. In this process, the baghouse filter cake is slurried into water, where fly ash and other insolubles are withdrawn and disposed of. The slurry is then pumped to another tank, where waste sulfuric acid (from virtually any source) and ferric ion are added. This tank is kept at an acidic pH (less than about 4) and a temperature of 120°F-150°F. The resultant double salts are granular, easily filterable precipitates that are stable and insoluble in water. These precipitates can be disposed of in open landfills without the need for clay-insulated cells.

Estimates of waste-disposal costs are given in Ref. 15. For the first technique discussed above, the annual operating cost was estimated to be approximately $7.40/ton of waste. Techniques used to insolubilize the baghouse filter cake could increase this cost by 50-100% or more.

3.2.2.2 Regeneration

Although the current thinking at utilities is along the lines of a throwaway cycle, the lack of availability and subsequent high cost of nahcolite may dictate that a regeneration process should be developed. Such a process would also greatly reduce environmental concerns about leaching of the sodium-salt wastes into subsurface water tables. One possible method of nahcolite regeneration is outlined in Ref. 16. With this process, the basic FGD system would be unchanged through the point where the spent cake is removed from the bag walls. Rather than conveying the spent cake to the disposal bins, it would be moved to a treatment plant. In the first step towards regeneration, the cake (containing sodium sulfites and sulfates, fly ash, and unreacted Na_2CO_3 or $NaHCO_3$) is put into a leaching tank, where the sodium salts are dissolved. The settled and filtered cake material, which would contain about 20% moisture, is disposed of as waste. To minimize the loss of sodium salts, the cake is washed before the final filtering operation.

The solution from the leaching tank is then pumped to another tank, where slaked lime [$Ca(OH)_2$] is added. The chemical reactions which occur in this stage of the process are as follows:

$$Na_2SO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_3 \tag{3.5}$$

$$Na_2CO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaCO_3 \tag{3.6}$$

$$NaHCO_3 + Ca(OH)_2 \rightarrow NaOH + CaCO_3 + H_2O \tag{3.7}$$

The calcium salts produced above settle out of solution and, along with some unreacted $Ca(OH)_2$, are filtered, washed, and disposed of as waste. The filtered washings are pumped to an evaporator where they are concentrated and then fed to an absorption tower. This tower is a counter-flow device in which the upflowing gas train is comprised of air and CO_2. In this tower, the following reactions take place:

$$NaOH + CO_2 \rightarrow NaHCO_3 \tag{3.8}$$

$$2NaOH + CO_2 \rightarrow Na_2CO_3 + H_2O \tag{3.9}$$

A crystallizer is the next stage of the process, and there $NaHCO_3$ and $Na_2CO_3 \cdot 10\ H_2O$ are produced. The adhering liquor is removed by washing, centrifuging, and drying. Crystals of $NaHCO_3$ and Na_2CO_3 are formed at this stage. These crystals are then stored for use in the system as regenerated reactant.

Data presented in Ref. 16 show that a plant burning low-sulfur coal, having a nahcolite/baghouse FGD system operating with a reagent stoichiometry of 1, and removing 60% of the stack-gas SO_2 would annually require 160 tons of nahcolite/MW without regeneration. With the regeneration process discussed above, the annual nahcolite requirements would be slashed to 1.26 ton/MW. The lime required in the regeneration process would be 96.2 ton/MW.

3.3 ECONOMIC AND MARKET FACTORS

3.3.1 Operating and Capital Costs

Reference 14 includes cost estimates for the southwest and midwest plants noted in Section 3.1.3. The original estimates were made for April 1973 dollars, but the figures have been adjusted to approximate BOY 1980 values.

For the southwestern plant, the basic capital investment for the components of the FGD system was estimated to be approximately $24,400,000. This figure includes a 15% contingency factor and a 3% contractor fee. When capital costs are included for crushing and grinding the nahcolite ore, waste-disposal land cost, and road construction, the overall capital investment associated with the FGD system becomes $27,200,000 ($34/kW).

For the midwestern plant, the investment for the components (including the contingency factor and contractor fee) was estimated at $68,500,000. When the capital costs of crushing and grinding the raw ore and the waste disposal site are included, the overall investment becomes $77,600,000 ($46/kW).

Maintenance costs for these plants consist chiefly of bag replacement: \$43 for the bag and \$8.60 for installation labor are representative of BOY 1980 costs. The expected lifetime of an individual bag was taken as two years. Maintenance in the rest of the system was taken to be 5% of the equipment cost exclusive of the baghouse. The power requirements were estimated according to such basic parameters as those presented in Table 3.1. Power costs of 5.2 mill/kWh in the Southwest and 9.5 mill/kWh in the Midwest were used in the estimates. The delivered price of the nahcolite ore to the southwestern plant was estimated to be \$40.90/ton, while that for the midwestern plant was \$32.60/ton. The operating costs for crushing and grinding the ore to the required size were \$1.50/ton and \$0.82/ton for the southwestern and midwestern plants, respectively.

A summary of the capital costs and O & M costs is presented in Table 3.3. It can be seen that the fixed O & M costs are expected to be quite low for each of the dry nahcolite systems. Variable O & M costs for the midwestern plant would be significantly greater -- this is due to the much greater quantity of nahcolite ore needed to remove 75% of the SO_2 produced by burning coal of 3.5% sulfur as compared to removing 70% of the SO_2 from coal containing 0.7% sulfur.

For purposes of comparison, Table 3.3 also shows cost estimates for a limestone wet-scrubbing system at each of the two locations. The O & M costs for the limestone systems were not divided into the two components. A total value was given based on a 68% capacity factor for the southwestern plant and a 65% factor for the midwestern plant. The total O & M cost for the dry nahcolite system can be seen to be 80-90% greater than the corresponding value for the reference limestone technology.

Table 3.3 Cost Estimates for Two Dry Injection/Baghouse Filtering Applications (BOY 1980 \$)

	Size of Expenditure			
	Southwestern Plant (One 800-MW Boiler)		Midwestern Plant (Two 840-MW Boilers)	
Required Expenditure	Dry Nahcolite	Wet Limestone	Dry Nahcolite	Wet Limestone
Capital investment				
Raw total (10^6 \$)	27.2	44.3	77.6	114.6
Per unit of plant output (\$/kW-yr)	34	55	46	68
Fixed O & M cost (\$/kW)	0.6	--	0.6	--
Variable O & M cost (mill/kWh)	1.9	--	4.7	--
Total O & M cost (mill/kWh)[a]	2	1.1	4.8	2.5

[a] Assuming capacity factors of 68% and 65% for the southwestern and midwestern plants, respectively.

Adapted from Ref. 14.

Within the accuracy of the estimates, the capital investment for a dry injection/baghouse filtering system is smaller than that for a wet limestone scrubber by 38% and 32% for the southwestern and midwestern applications, respectively. This factor exemplifies the relative simplicity of the nahcolite system compared to that of wet limestone FGD, and this has been purported to be an advantage of a dry system. For the southwestern plant using low-sulfur coal, the operating costs (including annual capital and O & M expenditures) for the nahcolite and the limestone systems are comparable. For the midwestern plants using coal with a considerably greater sulfur content, the annual cost of the limestone scrubber system is somewhat lower than the cost of the dry nahcolite process.

Reference 15 also presents cost estimates for a coal-burning power plant utilizing a nahcolite/baghouse filtering process. The reference conditions for this study include a new 500-MW boiler burning western coal composed of 1% sulfur and 10% fly ash and with a heat value of 10,500 Btu/lb. This coal would be burned at a rate of $4.3 \cdot 10^5$ lb/h. The flue gas would exit the boiler at 890°F and a flow rate of $8.9 \cdot 10^5$ scfm ($4.3 \cdot 10^6$ lb/h). After leaving the boiler and flowing through an economizer, the gas would flow through an air preheater for cooling to 400°F. The gas then goes to the baghouse, where the SO_2 and fly ash are removed. After leaving the baghouse, the gas goes through another air preheater for cooling to 225°F. The gas is then discharged up the stack. The rationale behind the two-stage air preheater will be discussed in paragraphs to follow. For the study reported in Ref. 15, it was assumed that 70% SO_2 removal was required. From the data in Fig. 3.2 it appears that a stoichiometric ratio of 1 could be used. This value was assumed in making the cost estimates. The plant was assumed to be located 750 mi from the nahcolite supplies and in a semiarid region with the water table 50 ft below the ground surface. The nahcolite-ore assay was 70%. The annual capacity factor was taken to be 80%.

Other factors affecting the cost of an FGD system can be determined from the parameters already defined and from various tests which have been conducted. For example, from the parameters already identified, the sorbent requirement can be determined to be 30,000 lb/h. The air-to-cloth ratio will affect several factors, the most important of which is the pressure-drop requirement. A greater air-to-cloth ratio will result in an increased differential pressure across the bag surface. The increased pressure drop will necessitate both higher capital investment and operating costs. An air-to-cloth ratio of 3 was chosen for the base case of this study. This value is within the range of values considered optimum for power plants fired with pulverized coal. The nahcolite required for precoating the filter bags was determined to be 6000 lb/h, with the remaining 24,000 lb/h injected into the flue-gas stream just downstream from the first-stage air preheater.

As noted earlier, the base case of this economic assessment is a new power plant. If the nahcolite FGD were to be retrofitted on an existing plant, several parameters would be different. A significant difference in the estimates would be apparent in the amount of nahcolite required to remove a given percentage of the SO_2. This difference would result from the fact that existing air preheaters are designed so that the temperature of the flue gas exiting the preheater is about 275°F. Referring back to Figs. 3.2 and 3.3, it is apparent that the stoichiometric ratio would have to be about 1.7 to yield 70% SO_2 removal. Thus, the sorbent cost and the waste-disposal costs would be significantly increased in a retrofitted application.

With the base power plant and FGD system described above, the capital investment and O & M costs were estimated. The results, originally expressed in 1977 dollars, are presented below as adjusted BOY 1980 values.

The total capital investment associated with the FGD system on the base case plant was estimated at $25,200,000. This investment is equal to $50/kW. Of this figure, approximately half is associated with the baghouse itself, and 26% is associated with the nahcolite-unloading section of the plant.

In making the operating cost estimates, the assumptions used include a price for nahcolite ore of $39/ton (including 1¢/ton-mi transportation costs); electricity cost of 2.2¢/kWh to operate equipment; and 9 man-yr per year of labor to operate the facility. Maintenance costs were set at $351,000/yr for the baghouse and at 4% of the capital investment for the remainder of the FGD system. The disposal costs for the spent nahcolite and fly ash were estimated at about $7.40/ton. In assigning a cost to the FGD system, a value of $7.15/ton for recovered fly ash was subtracted from the total costs because this material would be recovered and disposed of even if an FGD system were not used.

With these assumptions as to the O & M costs, the fixed and variable O & M costs for this new 500-MW plant were estimated at $2.7/kW-yr and 1.6 mill/kWh, respectively. A 70% annual capacity factor for this plant would make the total O & M cost equal to approximately 2 mill/kWh. An annual fixed-charge rate of 15% would make the total operating cost of the dry nahcolite FGD system equal to 3.2 mill/kWh (again assuming a 70% capacity factor).

As a point of comparison, cost estimates were also made for the same 500-MW plant equipped with a limestone-slurry FGD scrubbing system. The overall capital investment was estimated to be $71,900,000 ($144/kW). The largest cost single item in the limestone system is for fly ash particulate removal, which represents 48% of the total investment.

The fixed and variable O & M costs for the limestone FGD system were estimated at $7.8/kW-yr and 0.8 mill/kWh. A 70% annual capacity factor for this plant would make the total O & M cost equal to approximately 2.1 mill/kWh. The total annual operating cost (including capital and O & M costs) calculated under the same assumptions as above for the dry nahcolite system is equal to 5.6 mill/kWh.

Because of the need to make various assumptions for specific parameters in the base-case study, it may be desirable to conduct a sensitivity analysis for some of these factors. Two parameters of interest are the sulfur content in the coal and the heating value of the coal. The curves shown in Figs. 3.4 through 3.7 indicate the changes in investment costs and annual costs that might be anticipated because of changes in these parameters. (The ordinate in these figures is 1977 dollars.) Three separate curves are shown in each figure, with flue-gas temperature as the parameter. For a retrofit FGD system as opposed to the new power plant assumed in the base case, the curves for the 275°F gas temperature can be used. This figure represents a typical gas temperature at the exit of the air preheater in an existing power plant.

The two most capital-intensive items in the nahcolite FGD system were seen to be the baghouse and the nahcolite-unloading section. For the

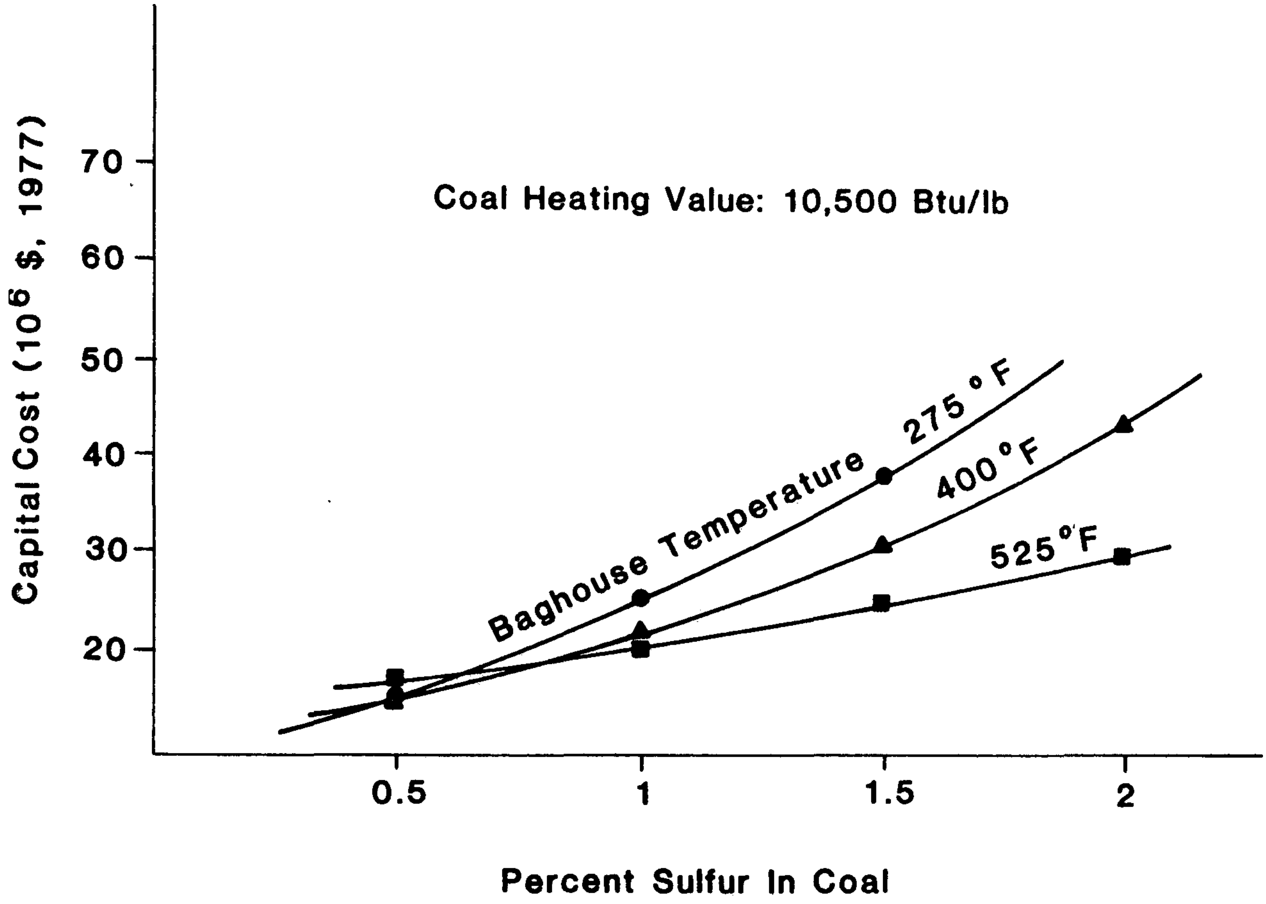


Fig. 3.4 Capital Cost of Dry Injection/Baghouse Filtering as Determined by Coal Sulfur Content and Baghouse Temperature (Source: Ref. 15)

baghouse, a curve relating cost per cfm versus the total gas flow rate was used in making the estimates. For the nahcolite-unloading section, it was assumed that the capital cost varied linearly with the weight of nahcolite utilized. Thus, this component would significantly increase the capital investment for either an increase in the sulfur content of the coal, or a decrease in the coal heat value for a fixed sulfur content. The affect of lowered temperature on the required nahcolite (and thus the cost of the nahcolite-unloading section) has been previously discussed.

The estimates presented in this section indicate that the capital cost for a dry nahcolite injection/baghouse filtering system is much less than that for an equivalent wet-limestone scrubbing system. A wide range of cost differentials was reported, however, with a low of 35% (ratio of nahcolite-to-limestone capital cost) and a high of 68%. This wide range is indicative of the difficulties arising when cost estimates are made for an immature technology.

For low-sulfur coal applications, the operation and maintenance costs for a dry-nahcolite system appear to be comparable to or slightly greater than those for an equivalent wet-limestone system. The lower capital costs make the dry nahcolite system an economically viable alternative to

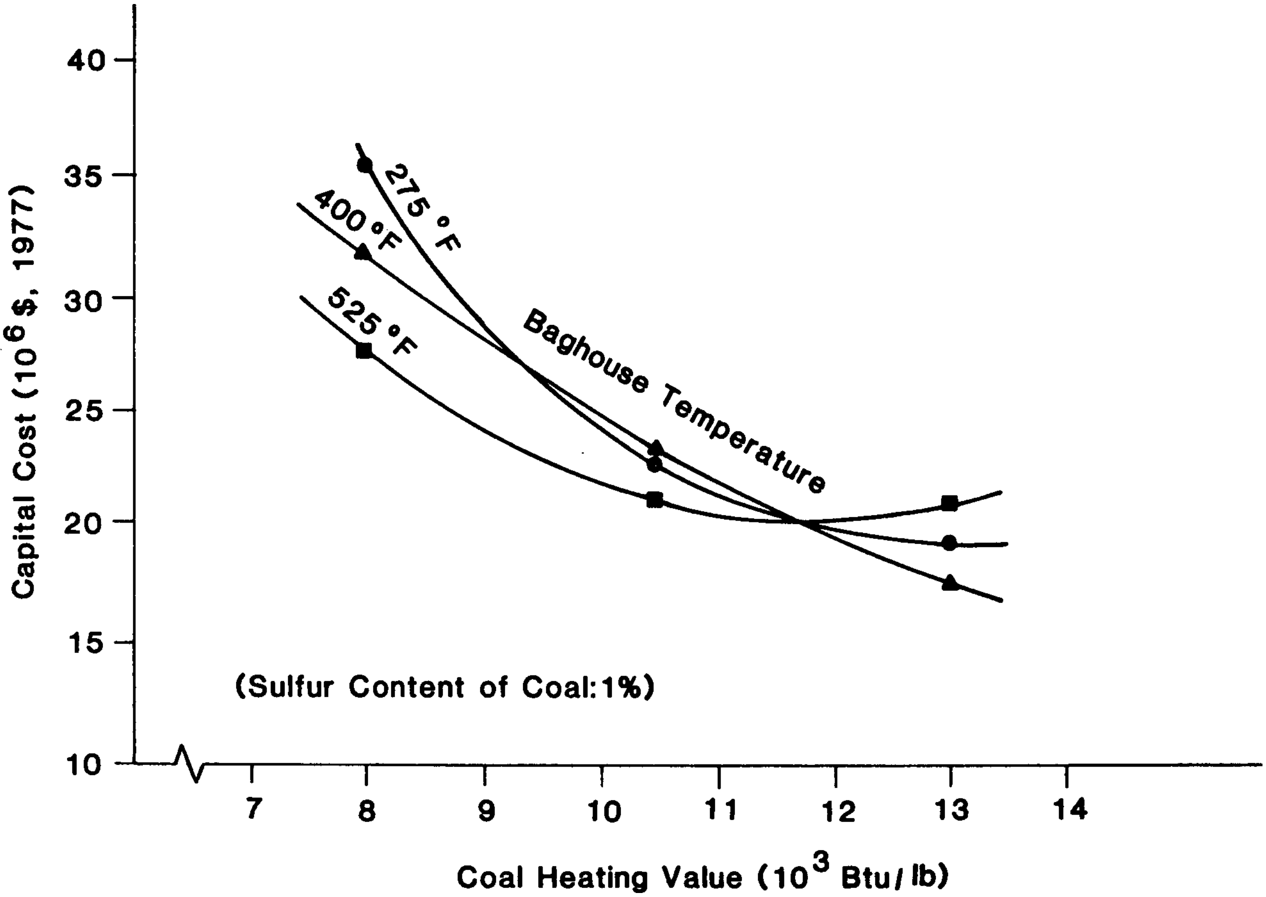


Fig. 3.5 Capital Cost of Dry Injection/Baghouse Filtering as Determined by Coal Heating Value and Baghouse Temperature (Source: Ref. 15)

wet-limestone scrubbers for low-sulfur coals. For high-sulfur coal, the high cost of the nahcolite ore would appear to make the dry injection/baghouse filtering process more costly than the reference limestone technology.

3.3.2 Potential Applications

The current evidence indicates that dry injection/baghouse filtering may be limited to low-sulfur coal. If SO_2 removal efficiency is in the range of 70-80%, this factor alone may limit the application of the process to low-sulfur coal even though the basic technology is applicable to all coals.

3.3.3 Market Potential

Two factors appear to be central in determining the market potential of this technology: 1) the potential limitation of its applicability to low-sulfur coals, and 2) the availability of sufficient nahcolite ore at a reasonable cost.

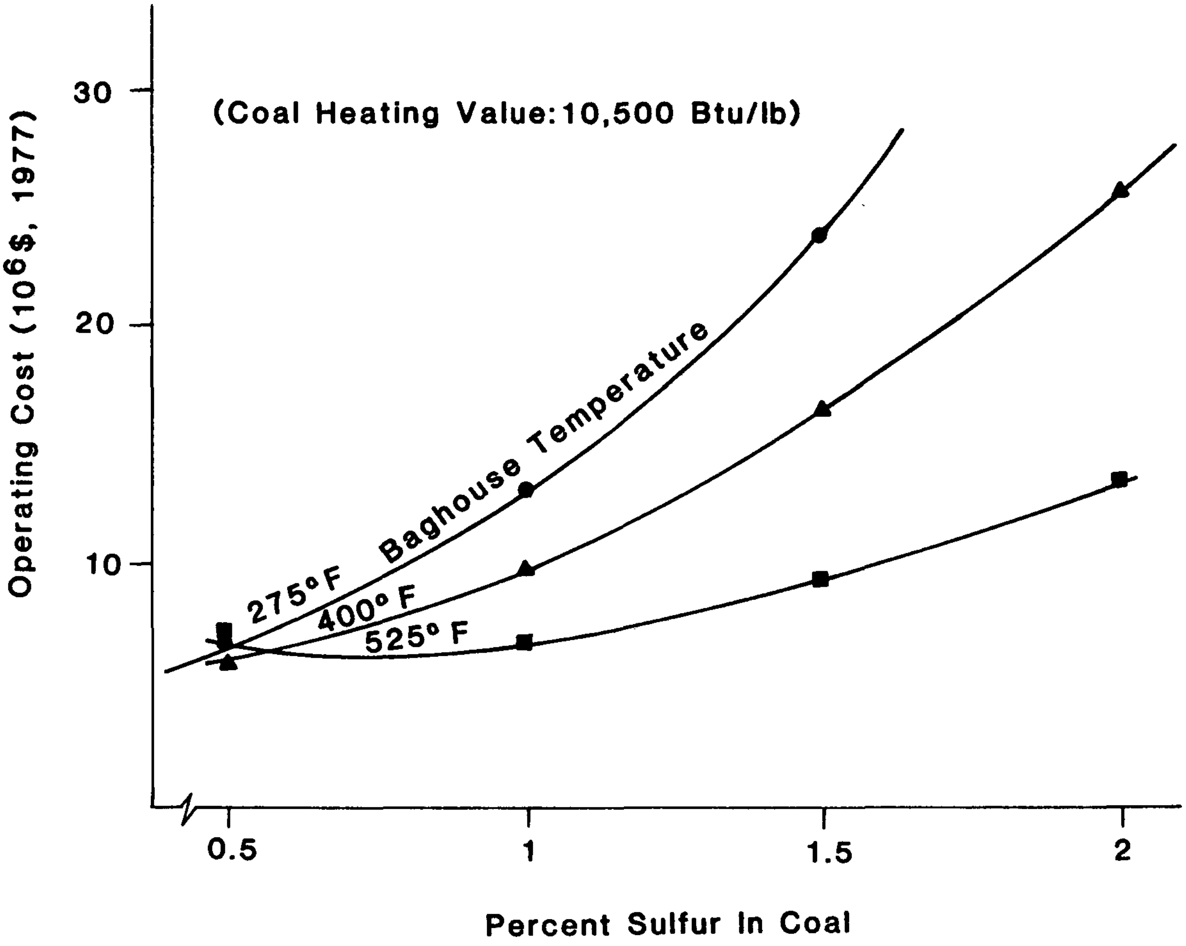


Fig. 3.6 Operating Costs of Dry Injection/Baghouse Filtering as Determined by Coal Sulfur Content and Baghouse Temperature (Source: Ref. 15)

3.4 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT

Current and planned test programs with baghouse dry sorbent include the following two efforts. A program has commenced at an 80-MW generating system owned by the City of Colorado Springs, Colo. This effort is being conducted by the Particulate Technology Branch (PaTB) of the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency (EPA). A 0.5%-sulfur western coal is used in this system. Currently, the effort consists of a pilot-sized baghouse (1000-1500 cfm) with a typical air-to-cloth ratio of 2/1. The principal objective is to evaluate the performance of the baghouse and the injected sorbent. Nahcolite and trona (naturally occurring Na_2CO_3) are being examined in this effort, which is to cover a 15-month period. An option exists to extend the pilot program or to construct a full-scale (80-MW) baghouse.

The second PaTB program currently underway is at Kerr Industries in Concord, N.C. A 35,000-acfm baghouse is installed on each of two 60-MW boilers. Each boiler uses coal with a sulfur content of 0.7% to 0.8%. The air-to-cloth ratio of these units can be varied from 1/1 to 3/1, with 2/1

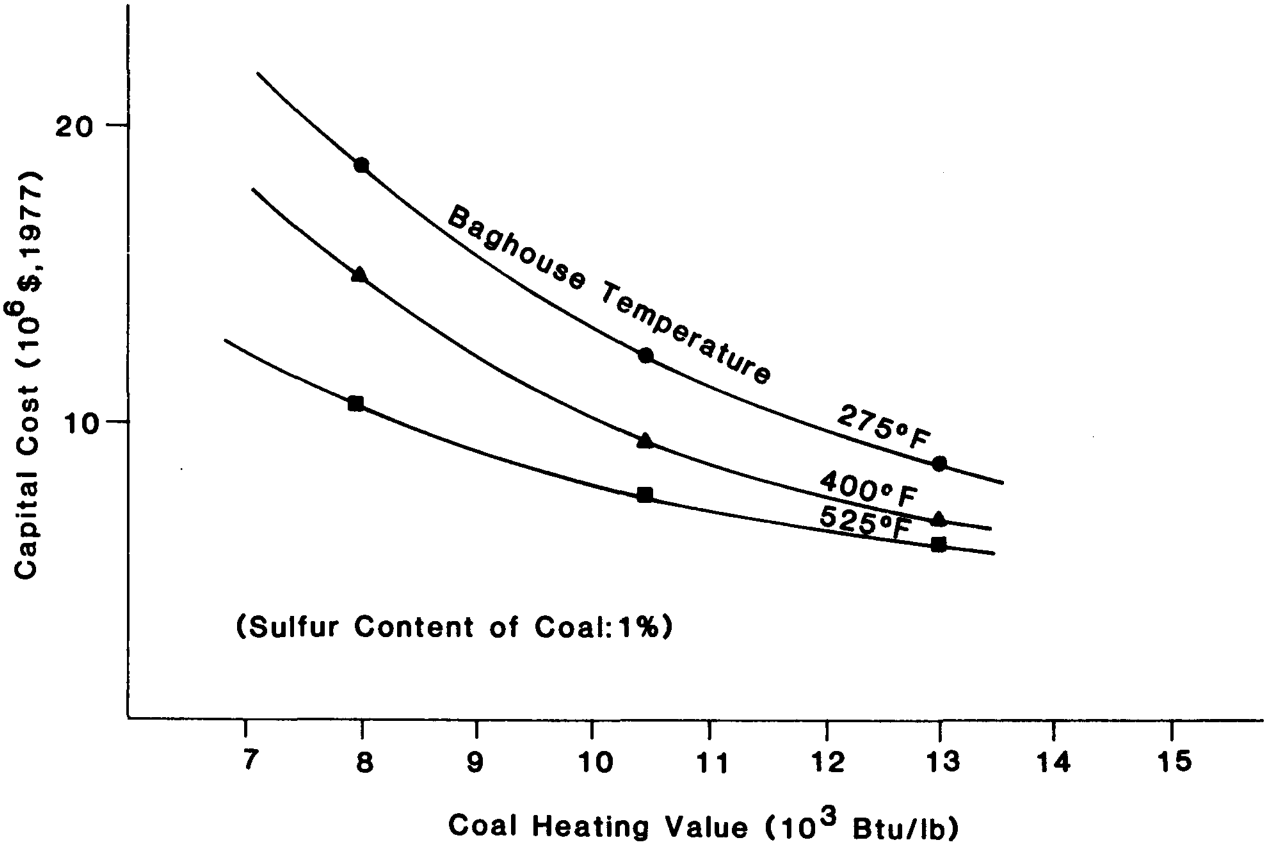


Fig. 3.7 Operating Costs of Dry Injection/Baghouse Filtering as Determined by Coal Heating Value and Baghouse Temperature (Source: Ref. 15)

being a typical operating value. The test program was to have begun in 1979 and a number of sorbents tested over a range of operating conditions. A regenerable-sorbent process is being considered as part of this test effort.

3.5 INDUSTRIAL-BOILER APPLICATIONS

Conceptually, the dry injection/baghouse filtering process is applicable to industrial as well as utility boilers. In fact, one of the test programs noted in Sec. 3.5 above is at Kerr Industries. The limited availability of nahcolite and the potential economic limitation to low-sulfur coals will impede the development of this process for industrial application in the same way as it will for utility boilers.

No information on cost estimates for industrial boilers has been obtained.

4 SPRAY DRYING/BAGHOUSE FILTERING

SUMMARY

The spray drying/baghouse filtering FGD technology represents an improvement over the dry injection/baghouse filter process in that sorbents other than the scarce nahcolite can be used; a somewhat higher SO_2 removal efficiency can be achieved; and, depending upon the sorbent used, the waste-disposal difficulties can be significantly reduced. In principle, this technology can be used with all types of coal; economic considerations, however, may limit its application to coals with a sulfur content of less than 2.5-3%. The technology has been commercially applied in both utility and industrial facilities.

4.1 TECHNICAL ASPECTS

4.1.1 Process Description

The concept of the two-stage FGD system evolved from various experiences with dry SO_2 removal technology. As discussed earlier in this report, several test programs have indicated favorable SO_2 removal with dry injection of nahcolite into the flue gas just upstream from a baghouse containing fabric filters. Tests of this type also showed that nahcolite or commercial sodium bicarbonate were the sorbents showing by far the greatest promise. As indicated in Ref. 17, however, by the spring of 1977 the three potential suppliers capable of providing nahcolite in the quantities needed for a reasonable application of this technolgy were experiencing difficulties in meeting government requirements which must be fulfilled prior to starting excavation of the Piceanie Creek Basin in northern Colorado. Because of the uncertainty in the availability of nahocolite (as well as in the cost, even if availability could be assured), various alternatives to dry nahcolite injection were examined. One such alternative is the two-stage spray drying/fabric filtering process.

The spray drying/fabric filtering process combines the capabilities of two proven technologies. It is a two-stage process. The first stage consists of a spray dryer contactor and provides the majority of the SO_2 removal. The second stage consists of a fabric-filter particulate collector (or baghouse), which collects fly ash and the dry products from the spray dryer as well as providing some secondary SO_2 removal. A simplified flow diagram for the two-stage process is shown in Fig. 4.1. As indicated in the figure, several materials can be used as the sorbent in this process. Similar basic techniques are used, however, for each sorbent.

With this process the sorbent is ground to a very fine powder and then mixed with water to form a solution, or slurry. The sorbent/water mixture is then sprayed as a fine mist into direct contact with the SO_2-laden flue gas. The heat in the gas evaporates the water, thereby leaving the powdered sorbent to react with the SO_2 in the gas and to form a dry salt. After

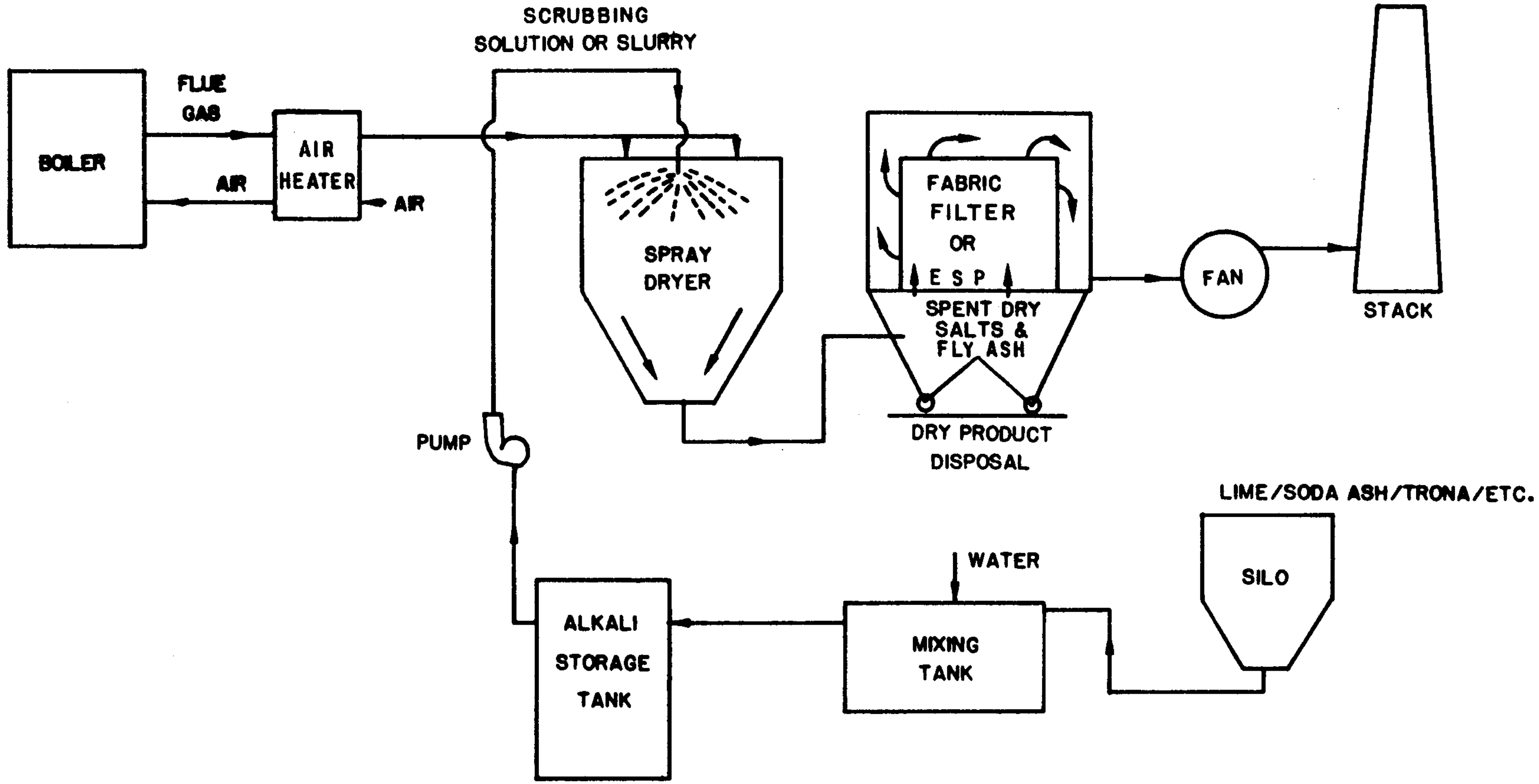


Fig. 4.1 Flow Diagram of Spray Drying/Baghouse Filtering FGD Process

leaving the sprayer, the gas flows into a fabric filter, where the dry salts and fly ash are collected. Some additional SO_2 removal also takes place in the fabric filters.

4.1.1.1 Test Programs

During the time of the nahcolite injection/fabric filtering tests at the Leland Olds plant, the potential nahcolite supply problems were becoming apparent. As a result, the test effort was expanded to incorporate a 7-ft-diameter spray dryer that would operate upstream of the fabric filters, which had already been tested. This phase of the test program lasted for about two months and included tests at flue-gas flow rates of 1000-4500 acfm and SO_2 concentrations of 400-2300 ppm. The sodium-carbonate solutions were supplied to the spray dryer at various flow rates so that the temperature drop across the spray dryer ranged from 90°F to 170°F. The operating temperature of the fabric filter ranged from 165°F to 230°F. Over this range of parameters no degradation of SO_2 removal efficiency was observed. Furthermore no serious equipment problems developed during a week-long period of continuous operation.

Some of the results obtained from the Leland Olds tests are shown in Table 4.1, where corresponding results with the dry nahcolite injection/fabric filter are also shown. The data in this table, which was taken from Ref. 17, were obtained with soda ash (Na_2CO_3) as the sorbent in the two-stage process. The temperature of the gas entering the fabric filter was approximately 200°F in the two-stage process, whereas it was about 290°F in the single-stage tests. With higher fabric-filter operating temperatures, the contribution of

Table 4.1 Comparative Effectiveness of Two Sorbents in the Leland Olds Tests of the Baghouse Process

	SO_2 Removal (%)			Sorbent Utilization (%)		
	Contribution of			Contribution of		
SR[a]	Spray Dryer	Fabric Filter	Total	Spray Dryer	Fabric Filter	Total
NAHCOLITE WITH FABRIC FILTER ONLY						
0.5	b	42	42	b	85	85
1	b	74	74	b	74	74
1.5	b	89	89	b	59	59
SODA ASH WITH SPRAY DRYER/FABRIC FILTER						
0.5	40	8	48	80	16	96
1	82	10	92	82	10	92
1.5	86	12	98	57	8	65

[a] Stoichiometric ratio.

[b] Not applicable.

Source: Ref. 17.

the fabric filter to the overall SO_2 removal of the fabric filter increased to values greater than 20% in the two-stage process.

When spray drying precedes the filtering phase of the baghouse process, improvements in both SO_2 removal and sorbent utilization are apparent, as Table 4.1 indicates. Thus, a viable alternative to dry nahcolite injection is indicated by these tests. An additional finding from the Leland Olds test program is that the capital investment for the two-stage process is only slightly greater than that for the single-stage fabric-filter process. An important factor in determining these costs is the flue-gas temperature in the baghouse. The lower temperature associated with the two-stage process allows the use of polyester fiber rather than the fiberglass required by the single-stage process. Furthermore, the size of the baghouse can be significantly reduced in the two-stage application, thus yielding additional savings. For example, the number of bags required at the 410-MW lignite-fired Coyote station was reduced by about 17% when a two-stage process was designed to replace the single-stage dry-injection process.

In addition to the tests conducted with soda ash, some tests have also been conducted with lime as the sorbent material. Some of the results of the lime test program are shown in Table 4.2 (again, from Ref. 17). A comparison of the data presented in Table 4.2 with the data presented earlier shows that the lime system is not as attractive as the soda-ash system either with regard to SO_2 removal or to sorbent utilization. However, on a cost per ton of SO_2 removed, the lime system is quite competitive with the soda-ash technique. This feature will be discussed further in a later section of this report.

4.1.1.2 System Flow Diagrams

The two-stage spray drying/baghouse filtering process has developed sufficiently so that several units have been contracted for and are currently under design or construction. Three of the earliest such units are for power plants burning North Dakota lignite coal. A brief description of the three units is provided below. These data provide some information on the material balances for this technology.

Table 4.2 Effectiveness of Lime As the Sorbent in a Two-Stage Baghouse Process

		SO_2 Removal (%)			Sorbent Utilization (%)		
		Contribution of			Contribution of		
SO_2 Concentration (ppm)	SR[a]	Spray Dryer	Fabric Filter	Total	Spray Dryer	Fabric Filter	Total
1635	1.21	50	25	75	41	21	62
1240	0.94	50	19	69	53	21	74
840	0.66	35	18	53	53	38	81

[a] Stoichiometric ratio.

Source: Ref. 17.

4.1.1.3 Coyote Station - Unit 1

The conditions existing at this new 410-MW station were used as the basis for the test program conducted at the Leland Olds plant. Due to the successful demonstration of soda ash in the two-stage process at Leland Olds, it was decided that it would also be used as the alkali sorbent at Coyote station.

The Coyote station is located near Beulah, N.D., and is owned by the Otter Trail Power Co. and several other utilities. This plant is a duplicate of the Big Stone plant, currently in operation near Big Stone City, S.D., except for certain site-sensitive items, the particulate-removal equipment, and the flue-gas-desulfurization equipment. The plant will burn lignite coal having a heating value of 7046 Btu/lb, an average sulfur content of 0.78%, and an ash content of approximately 7%. Flue gas will leave the air preheater at 285°F and a flow rate of 1,890,000 acfm at full load. The FGD system includes four spray-dryer chambers arranged in parallel. Each chamber is sized for about one-half million acfm of gas and incorporates three centrifugal atomizers. A rotating (18,000 rpm) disk atomizes the water/alkali solution. The resultant mist is injected in a cross-current-flow pattern to the flue gas entering at the top of the chamber. A multicompartment fabric filter collects the dry salts and the fly ash. A combination mechanical and pneumatic cleaning cycle is used on the filters. A precipitator could have been used instead of the fabric filters, but the tests at Leland Olds indicated that from 12% to 20% SO_2 removal could be accomplished with the filters. Therefore, fabric filters were selected to realize both particulate removal and additional SO_2 removal. A minimum SO_2 removal of 70% is guaranteed, and soda ash utilization is expected to approach 100%. The waste products are collected in the baghouse hoppers for ultimate disposal in an adjacent mine supplying the coal. Although a throwaway disposal system is currently envisioned, the plant design does not preclude the use of a reclamation system for sulfur and soda ash if the economics warrant it. This plant is currently under construction and is scheduled for completion in June 1981.

4.1.1.4 Antelope Valley

This station has two 440-MW plants owned by Basin Electric Power Cooperative and is located northwest of Bismark, N.D. The alkali sorbent for the FGD system at Antelope Valley will be slaked lime. An underlying reason for using slaked lime was concern over potential disposal problems and the higher costs associated with sodium-based reagents. An extensive test program was therefore initiated to investigate the effectiveness of several reagents in a two-stage spray drying/baghouse filtering system. The design conditions for Antelope Valley provided the basic parameters used in the test effort. During the test program a great deal of experience was gained with important operating parameters such as slurry feed rates, atomizer design and speeds, spraydown temperatures, and many others. This experience allowed the stoichiometric ratio to be improved so that nearly 100% utilization of the lime feed was attained. At Antelope Valley the lime feed rate will be essentially stoichiometric. This low ratio is possible because of the alkalinity available in the fly ash can be used to supplement the slaked lime. The fly ash content in the coal burned at Antelope Valley averages about 7.8%.

To meet the requirements of the North Dakota State Department of Health, the SO_2 and particulate removal system is designed to operate at 62% SO_2 removal for performance lignite (average sulfur content) and 78% removal for design lignite (maximum sulfur content).[18,19] These removal efficiencies are required to meet the overall limitation of 0.78 lb $SO_2/10^6$ Btu.

The FGD system will incorporate five spray dryers (four in operation and one spare), and each of them will utilize one rotary atomizer. A baghouse of 28 compartments will be used to collect particulate matter and provide additional SO_2 removal. About 8000 individual bags will be in the baghouse. Each bag is 1 ft in diameter and 35 ft high. All bags will be made of fiberglass. An air-to-cloth ratio of 2.19 will be achieved under the maximum operating conditions. Reverse air cleaning will be used to remove collected salts and particulate matter from the bags.

The waste product removed from the baghouse is a very fine grain powder-like material similar in particulate-size distribution to the fly ash normally removed from a coal-fired plant. For Antelope Valley the method of waste disposal will be to transport it to depleted regions of the mine for use as landfill. It is believed that there are no apparent disadvantages in disposing of a calcium-based dry waste product over and above those associated with disposing of fly ash. This may not be the case with a sodium-based waste product. To illustrate this concern, it is noted in Ref. 19 that the solubility of a calcium-based product is 3-7%, whereas sodium-based waste products have been found to be 50-60% soluble. The waste product from Antelope Valley is fairly cementitious and impermeable, but potential problems may arise due to weathering, erosion, dust suppression, and structural stability. More specific disposal procedures will be developed as the product becomes available. Commercial operation is scheduled to commence in April 1982.

Each unit at Antelope Valley has a gross generating capacity of 440 MW and a net of 385 MW. Lignite coal with an average sulfur content of 0.68% and an average heat value of 6600 Btu/lb will be burned at a full power rate of about 375 ton/hr. Table 4.3 gives the flue-gas characteristics expected at Antelope Valley for the average fuel-analysis and full-operating-power conditions. The "inlet" column refers to the inlet of the spray dryer and the "outlet" column refers to the outlet of the baghouse.

4.1.1.5 Laramie River - Unit 3

The Laramie River power station is also owned by the Basin Electric Power Cooperative. The first two units at this station are equipped with conventional wet scrubbers; but Unit 3, scheduled for operation in April 1982, will be fitted with a two-stage system of spray drying and baghouse filtering. This system will be a variation of the one at Antelope Valley in that four "reactors" (three in operation and one spare) will be used and each reactor will be followed by an electrostatic precipitator (ESP). Each reactor is equipped with 12 Y-shaped jet nozzles. The dual-fluid atomizers will use steam as the atomizing fluid and lime slurry as the reagent. Three hoppers will be located under each reactor chamber to collect some of the waste products before the flue gas goes into the precipitator. About 3% of the flue gas will bypass the air preheater and the spray dryers and will enter the reactor discharge plenum, where it will reheat the flue gas discharged from

Table 4.3 Flue-Gas Characteristics of Spray Dryer/ Baghouse Filter System at Antelope Valley

Characteristic	At Inlet[a]	At Outlet[b]
Gas flow (acfm)	2,055,000	1,894,380
(scfm)	1,248,900	1,369,000
Gas temperature (°F)	310	185
Gas mass flow (lb/h)	5,690,000	6,097,000
Moisture (% of volume)	15.6	19.4
SO_2 content (ppm by volume)	800	304
(lb/h)	10,120	3,845
(lb/10^6 Btu)	2.07	0.78
Particulates (lb/h)	88,070	210
(g/acf)	5	0.012
(g/scf)	7.7	0.018

[a] Of spray dryer.

[b] Of baghouse.

Source: Ref. 18.

the spray dryers. The reheated flue gas will then go to an ESP of the same design and size as those on Units 1 and 2. To meet the North Dakota environmental standards, the Laramie River FGD system has been designed to remove 85% of the SO_2 in the gas generated by 0.54%-sulfur coal, and to remove 90% of the SO_2 generated by 0.81%-sulfur coal.

The gross electrical output from Laramie River - Unit 3 is 575 MW, and the net is 500 MW. The coal burned will have an average sulfur content of 0.54% and an average heating value of 8139 Btu/lb. A coal burn rate of 340 ton/hr will be required to produce the rated power.

Table 4.4 gives the flue-gas characteristics for Laramie River - Unit 3. The "inlet" column refers to the inlet of the spray dryers and the "outlet" column to the outlet of the electrostatic precipitator. As indicated in this table, the North Dakota requirement of less than 0.78 lb $SO_2/10^6$ Btu is expected to be easily attained by the Laramie River unit. (See Table 4.4.)

4.1.2 Process Chemistry

The chemical reactions which occur with this process depend, of course, on the sorbent used. When lime is used as the sorbent, the basic chemical reactions are similar to those which take place in a wet lime-based scrubber except that the resultant salts are dry. The mechanism for these reactions is as follows.

The SO_2 and the CO_2 in the flue gas are dissociated in the alkaline medium in accordance with these reactions:

Table 4.4 Flue-Gas Characteristics of Spray Dryer/ Electrostatic Precipitator at Laramie River

Characteristic	At Inlet[a]	At Outlet[b]
Gas flow (acfm)	2,300,000	1,946,000
(scfm)	1,320,000	1,405,000
Gas temperature (°F)	286	157
Gas mass flow (lb/h)	6,188,000	6,461,000
Moisture (% of volume)	11	17
SO_2 content (ppm by volume)	530	80
(lb/h)	7,320	1,100
(lb/10^6 Btu)	1.33	0.2
Particulates (lb/h)	78,860	300
(g/acf)	4	0.015
(g/scf)	7	0.02

[a] Of spray dryer.

[b] Of electrostatic precipitator.

Source: Ref. 19.

$$SO_2 + H_2O \rightleftarrows H_2SO_3 \rightleftarrows H^+ + HSO_3^- \rightleftarrows 2H^+ + SO_3^= \qquad (4.1)$$

$$CO_2 + H_2O \rightleftarrows H_2CO_3 \rightleftarrows H^+ + HCO_3^- \rightleftarrows 2H^+ + CO_3^= \qquad (4.2)$$

$$SO_2 + H_2O + SO_3^= \rightleftarrows 2HSO_3^- \qquad (4.3)$$

$$SO_2 + HCO_3^- \rightleftarrows HSO_3^- + CO_2 \qquad (4.4)$$

The solids are also dissociated via the reactions:

$$CaCO_3 + H^+ + HCO_3^- \rightleftarrows Ca^{++} + 2H^+ + 2CO_3^= \qquad (4.5)$$

$$Ca(OH)_2 \rightleftarrows Ca^{++} + 2O\bar{H} \qquad (4.6)$$

Insoluble sulfites and sulfates are then formed.

$$Ca^{++} + SO_3^= + 1/2\ H_2O \rightleftarrows CaSO_3 \cdot 1/2\ H_2O \qquad (4.7)$$

$$Ca^{++} + CO_3^= \rightleftarrows CaCO_3 \qquad (4.8)$$

$$CaSO_3 \cdot 1/2\ H_2O + 1/2\ O_2 + 3/2\ H_2O \rightleftarrows CaSO_4 \cdot 2H_2O \qquad (4.9)$$

These salts then precipitate from the solution as crystalline encrustations on the surfaces of the sorbent particles. The tests at Colorado Springs showed the importance of the sequence of reactions leading to the formation of $CaCO_3$. It is the calcium tied up in $CaCO_3$ that makes the partial-reaction-product recycle an effective way of increasing sorbent utilization.

When soda ash is used as the sorbent, the reactions are the same as when nahcolite is used as a dry sorbent except that the first step of decomposing sodium bicarbonate into sodium carbonate has been eliminated.

4.1.3 Process Performance

SO_2 Removal Efficiency. The plants described above are intended to remove up to 90% of the SO_2 generated by burning low-sulfur coal.

System Reliability. Whereas no specific numbers are available at this time, the reliability experienced in the test programs discussed above was sufficient so that, as of May 1980, 10 such FGD systems were on order for utilities.[20] The redundant spray dryer module incorporated into recent designs will improve the FGD system availability.

Sulfur Recovery. To date, this technology is viewed as a throwaway process.

Effect of Coal Type on Performance. The test efforts and the commercial applications for this system to date have used low-sulfur coal only. Performance parameters with medium- or high-sulfur coal have yet to be established. It is believed that the two-stage spray drying/baghouse filtering is capable of high SO_2 removal efficiencies in systems burning high-sulfur coal but that it may be too expensive to do so. Sulfur contents of 2.5-3% may represent a breakeven point between this two-stage dry process and conventional wet scrubbing.

Effects of Plant Size and Operating Mode on Performance. The units described above are designed to follow the output of the power plants on which they have been tested. Like the power plants, therefore, they have a turndown capability of 25%.

4.1.4 Water, Chemical, and Power Requirements

Water requirements for this two-stage process are very low, with an expected liquid-to-gas ratio of less than 0.3 gal/1000 scfm of flue gas. For the three units described above, the water requirement ranges from approximately 300-400 gal/min. Expressed in other terms, the water requirement is expected to be less than 1 gal/m per MW. Virtually all of this water may be obtained from cooling tower blowdown or other boiler-waste streams.

Estimates of sorbent requirements were made for the Antelope Valley and the Laramie River units. The lime requirement for Antelope Valley is estimated to be 0.006 ton/MWh, whereas the value for the Laramie River Station is 0.007 ton/MWh.

Reference 19 gives the FGD electrical power requirements for the Antelope Valley and Laramie River stations as 5726 and 2451 kW, respectively. The values represent less than 1% of the heat input to the boilers.

4.2 ENVIRONMENTAL IMPLICATIONS

The degree of environmental difficulty associated with this technology is largely dependent upon the sorbent used. For the calcium-based sorbents, it is believed that there are no FGD waste-disposal disadvantages beyond those normally associated with the disposal of fly ash. For the sodium-based sorbents, however, special landfill techniques, such as those associated with the nahcolite dry injection process, may be required.

4.3 ECONOMIC AND MARKET FACTORS

4.3.1 Operating and Capital Costs

Several estimates have been made of the costs of constructing and operating two-stage spray dryer/filter systems. In addition, some information is available for the contracted price of the three systems discussed earlier. These data are presented below. Where available, the estimated costs of a wet lime/limestone scrubber for the identical plant are also presented.

4.3.1.1 Coyote Station - Unit 1

Reference 21 includes estimates of the Coyote station investment and operating costs. These estimates include the complete turnkey installation from the air preheater outlet to the stack connections, exclusive of the induced-draft fans and the waste-removal system (which would be required without an FGD system). When updated to approximate BOY 1980 dollars, the capital investment for Coyote station is approximately $95/kW. Annual operating and maintenance costs (assuming a soda-ash price of $92/ton) were estimated at $7,980,000 (3 mill/kWh). This figure does not include the costs of waste disposal.

4.3.1.2 Antelope Valley

Reference 19 gives the capital investment costs and the expected operating costs for the FGD system at Antelope Valley. Also included in Ref. 19 are the estimated costs for installing and operating a wet limestone FGD system design to meet the same requirements as the dry system. This cost comparison is summarized in Table 4.5. An annual plant factor of 75% and the present worth of operation and maintenance cost as of the 1982 commercial date are used in the table.

As shown in this table, significant annual savings are expected by using the dry two-stage sprayer/filter process instead of the conventional wet limestone scrubber.

4.3.1.3 Laramie River - Unit 3

Cost figures for the Laramie River - Unit 3 FGD system are presented in Table 4.6. They are based on an annual plant factor of 75% and the present worth of operation and maintenance costs as of 1982.

Table 4.5 Antelope Valley Cost Comparisons

	Size of Expenditure	
Required Expenditure	Dry Scrubber	Wet Scrubber
Capital investment ($/kW)	113	127
Fixed O & M ($/kW-yr)	2.2	4.2
Variable O & M (mill/kWh)[a]	0.5	0.6
Total O & M (mill/kWh)[a]	0.8	1.2

[a]Does not include waste-disposal costs.

Source: Ref. 19.

Table 4.6 Laramie River Cost Comparisons

	Size of Expenditure	
Required Expenditure	Dry Scrubber	Wet Scrubber
Capital investment ($/kW)	100	121
Fixed O & M ($/kW-yr)	1.8	3.7
Variable O & M (mill/kWh)[a]	0.5	0.3
Total O & M (mill/kWh)[a]	0.8	0.9

[a]Does not include waste-disposal costs.

Source: Ref. 19.

Table 4.6 also indicates that savings can be realized with the dry two-stage sprayer/filter system as opposed to the conventional limestone wet scrubber. This savings occurs in spite of the local availability of limestone near the Laramie River site, which is reflected in the low variable O & M cost for the wet scrubber.

It was noted earlier that while lime was not as effective as soda ash in terms of SO_2 removal and sorbent utilization, it may have an advantage on a cost-per-ton basis of removing SO_2. This factor is summarized in Table 4.7, where estimates for a boiler burning lignite and with a heating value of 7046 Btu/lb are shown. In preparing this table, it was assumed that the delivered price of soda ash was $75/ton, while that for lime was $53/ton.

Expressed in the terms of Table 4.7, it is apparent that the SO_2 removal cost for the lime system increases with increasing sulfur content at a much faster rate than does the soda-ash system. These data indicate that any cost advantages that exist with the lime system disappear at a sulfur-content level of about 1.4% and that the more reactive soda ash would be more economically attractive. However, further investigation of this potential may lead to the demonstration that lime usage can be commercially attractive even in applications for coals of high-sulfur content.

Table 4.7 Soda Ash Vs. Lime As a Spray-Dried Sorbent: Comparative SO_2-Removal Costs, in Dollars and Sorbent Required

Sulfur Content (%)	Total Cost of Removing SO_2 ($/ton)		Sorbent Needed to Remove SO_2 (lb/lb)	
	Soda Ash	Lime	Soda Ash	Lime
0.8	130	86	1.74	1.62
1	138	103	1.84	1.94
1.2	146	129	1.95	2.43
1.4	155	172	2.07	3.24

4.3.1.4 Additional Economic Assessments

Economic assessments of spray-dryer FGD units are presented in Ref. 22 for three coal types: a low-sulfur western, a low-sulfur eastern, and a high-sulfur eastern coal. These costs are compared with a corresponding limestone wet scrubber. For the low-sulfur western coal application, both soda ash and lime spray dryer systems are examined. For the eastern coals, however, soda ash was not considered because of the great uncertainty in determining the economics associated with the disposal problems of soluble sodium wastes in high rainfall regions.

The FGD systems are designed to meet the 1979 New Source Performance Standards for a new 500-MWe boiler firing pulverized coal. The systems have one redundant train, 50% emergency flue-gas bypass, and are designed and costed on the basis of current technology. A waste-disposal site consists of a clay-lined pond for the soda-ash unit and landfills for the others. Waste recycle is used to reduce sorbent consumption in the lime-based spray-dryer application for low-sulfur western coal. This recycle allows the high alkalinity of this coal to be used advantageously.

Some of the design and economic parameters used in this assessment are shown in Table 4.8. The percentage of flue-gas bypass may be of special interest. For lime spray-dryer applications, the bypassed flue gas is used to reheat the gas that has passed through the spray dryer before it enters the fabric filters. This procedure ensures that the gas entering the fabric filters is dry, while allowing a closer approach to saturation to be achieved in the spray dryer and thus enhancing the SO_2 removal efficiency. For the two low-sulfur cases, the bypassed flue gas is at 300°F; it is at 700°F for the high-sulfur eastern coal case. This latter case necessitates that the flue gas is also bypassed around the air preheater, and this reduces the boiler efficiency somewhat. When soda ash is used in a spray dryer, the approach to saturation is not so important, and the need for reheat prior to the fabric filter is obviated. For the wet limestone scrubber system, the bypass is used to minimize reheat requirements prior to discharging the gas in

Table 4.8 Design and Economic Comparison of Spray-Drying and Wet-Scrubbing Systems for Three Types of Coal

	Low-Sulfur Western Coal			Low-Sulfur Eastern		High-Sulfur Eastern	
	Spray Drying						
Parameter	Using Soda Ash	Using Lime	Wet Scrubbing Using Limestone	Spray Drying Using Lime	Wet Scrubbing Using Limestone	Spray Drying Using Lime	Wet Scrubbing Using Limestone
Stoichiometric ratio	1	1.2	1.1	1.3	1.1	1.5	1.3
Sorbent cost (1984 $/ton)	145	102	8.5	102	8.5	102	8.5
Flue gas bypass (%)	0	22	27.9	19	25.2	4	0
Coal sulfur content (%)	0.7	0.7	0.7	0.7	0.7	3.5	3.5
Coal ash content (%)	9.7	9.7	9.7	16	16	16	16
Coal heat value (10^3 Btu/lb)	9.7	9.7	9.7	11.7	11.7	11.7	11.7
L/G[a] (gal/1000 acf)	0.13	0.2	80	0.3	80	0.3	80
Reheating needed[b] (°F)	0	0	0	0	10	0	43

[a] Liquid-to-gas ratio.

[b] In order to achieve a stack temperature of 175°F.

Source: Ref. 22.

the stack. For high-sulfur coal, however, the SO_2 removal requirement necessitates that all the flue gas be scrubbed (i.e., no bypass); this means the reheat must be achieved by other means.

Capital investment estimates for the various FGD plants examined in this section are shown in Table 4.9. These values, which are the midpoint of a range of ±4 \$/kW given in Ref. 22, are based on mid 1982 costs. The absolute accuracy is estimated at -20% to +40%, but cost comparisons are estimated to be within ±10%.

Several conclusions can be drawn from the data in Table 4.9. First, the soda-ash spray-drying system requires a slightly greater capital investment than the lime spray-drying system. This greater investment results from the higher construction and land costs associated with the clay-lined pond used for disposal of the sodium-based wastes. The capital cost comparisons with respect to the limestone wet-scrubbing system indicate that the differential between a spray dryer and a wet scrubber increases with sulfur content. This result occurs because the major cost area of the wet system is the SO_2-absorption component (representing more than one-third of the direct capital costs). This component's cost increases by about 50% as the sulfur content increases from 0.7% to 3.5%. On the other hand, the SO_2-absorption component of the spray-drying system is only about half that of the limestone wet-scrubbing system and increases only slightly with sulfur content.

The first-year and levelized annual revenue requirements were also estimated for these systems and are presented in Table 4.10. The first-year requirements include raw materials, operating and overhead costs, and the levelized capital charges. The levelized annual costs include a multiplier of 1.886 applied to direct operating expenses and represents an assumed 6% annual inflation rate and a 10% discount rate over the 30-yr lifetime. A capacity factor of 62.8% was assumed in all results shown in this table.

Table 4.9 Comparative Capital Investment of Spray-Drying and Wet-Scrubbing Systems for the Three Coals

FGD System	Raw Total (\$$10^6$)	Per Unit of Output (\$/kW)	Relative to Limestone Wet Scrubbing (%)
Low-sulfur western coal			
Spray drying w/soda ash	78	156	91
Spray drying w/lime	74	148	86
Wet scrubbing w/limestone	86	172	100
Low-sulfur eastern coal			
Spray drying w/lime	74	148	80
Wet scrubbing w/limestone	92	184	100
High-sulfur eastern coal			
Spray drying w/lime	92	184	77
Wet scrubbing w/limestone	120	240	100

Source: Ref. 22.

Table 4.10 Comparative Annual Revenue Requirements for Spray-Drying and Wet-Scrubbing Systems for the Three Coals

	Annual Revenue Requirements					
	First-Year Costs			Levelized Annual Costs		
FGD System	Raw Total ($\$10^6$)	Per Unit of Output (mill/kWh)	Relative to Limestone Wet Scrubbing (%)	Raw Total ($\$10^6$)	Per Unit of Output (mill/kWh)	Relative to Limestone Wet Scrubbing (%)
Low-sulfur western coal						
Spray drying w/soda ash	19.5	7	91	26.5	9.6	90
Spray drying w/lime	18.5	6.7	86	24.5	8.9	83
Wet scrubbing w/limestone	21.5	7.8	100	29.5	10.7	100
Low-sulfur eastern coal						
Spray drying w/lime	17.5	6.4	74	23.5	8.5	75
Wet scrubbing w/limestone	23.5	8.5	100	31.5	11.5	100
High-sulfur eastern coal						
Spray drying w/lime	27.5	10	85	40.5	14.7	89
Wet scrubbing w/limestone	32.5	11.8	100	45.5	16.6	100

Source: Ref. 22.

In the low-sulfur western coal application, the lime spray dryer system is seen to be less costly than the soda ash system. This difference is principally due to the greater price of soda ash and the high capital investment requirement. For the limestone scrubbing system, the sorbent costs are quite low, and the higher revenue requirements reflect the greater capital expenditure and greater maintenance costs. The latter are estimated to be more than double the maintenance costs of the spray-drying systems.

For the low-sulfur eastern coal applications, the economic competitiveness of the lime spray dryer increases with respect to the limestone wet scrubber. This relationship results partially from the lower cost of lime in the eastern region and partially from the higher capital costs of the limestone system in this region.

With high-sulfur eastern coal, the spray-drying system remains the lower cost alternative. There is a smaller economic advantage, however, for the spray-dryer system in a high-sulfur application as compared to a low-sulfur application. This results from the significantly increased quantity of sorbent required for high-sulfur coals. Because the sorbent cost comprises a greater fraction of the total cost of a spray-drying system than of a limestone system, the increased sorbent usage for high-sulfur coal causes a greater percentage increase in the total cost of the spray-drying system.

The results discussed above for the high-sulfur eastern coal were somewhat surprising because the general feeling of people in the industry is that spray dryer costs exceed those of a limestone wet scrubber. The differences reported in Ref. 22 resulted from reduced equipment costs and lower maintenance costs of the spray dryer relative to those of a wet scrubber. Higher sorbent costs for the spray dryer are expected, but the overall cost advantage still lies with the spray dryer. It must be recognized, however, that all the major tests of spray drying conducted thus far have been applied to the burning of low-sulfur coal -- and that the greatest uncertainties in spray-dryer performance and costs are therefore associated with the burning of high-sulfur coal.

Another economic assessment of spray-dryer FGD technology is discussed in Ref. 23. This reference reports the results obtained from a number of vendors who were asked to provide cost estimates for a spray dryer retrofitted to the boilers in two power stations of a northeastern utility. The FGD system would be required to allow the utility to burn low-sulfur (0.9%) coal from West Virginia with SO_x and particulate emissions controlled to the present uncontrolled levels produced by burning No. 6 oil.

Estimates were to be made for two separate power plants, designated A and B. Power plant A has two boilers; one with a net generating capacity of 335 MW and the other with one of 491 MW. The overall plant heat rate is 10,264 Btu/kWh. The two boilers have 24 and 34 years, respectively, of operating life remaining. Power plant B has a single 900-MW boiler with a heat rate of 10,265 Btu/kWh. This plant has 30 years of operating life remaining. To meet the current uncontrolled emission from burning No. 6 oil, the particulate emissions from each plant must be reduced by 99.5%. The SO_2 emissions must be controlled to 0.3 $lb/10^6$ Btu, which requires an 82% removal efficiency. The vendors were instructed to design system redundancy into the plants. Two of the designs had an extra spray dryer module, whereas the other two had provisions for an extra atomizer head in each of the spray dryer modules. Both the fabric filter and spray dryer modules were designed to allow one module to be shut down for maintenance while the other modules remained operational.

Total capital investment requirements were made by asking the vendors to estimate the direct capital costs. The estimates were then doubled to obtain the total (direct plus indirect) capital investment -- a factor of 2 has been developed by TVA analysts for this purpose. These estimates have an estimated accuracy of ±30%.

Estimates were made for two different locations -- one a generalized northeastern U.S. location and the other a highly urbanized area on the order of northern New Jersey-southern New York.

The resultant total capital cost estimates are shown in Table 4.11. It may be noted that three of the four vendors combined the gas flows of units 1 and 2 of Plant A into a single FGD system. The estimates in this table show a maximum variation of ±18% from the average for each plant in the generalized location. Each of the three vendors providing estimates for these plants was consistently at, above, or below the average value. A significant capital cost increase would be expected for a plant in a highly urbanized area, and increases of 40-50% were estimated by vendor 3 for that situation.

Table 4.11 Capital Cost Estimates for Spray Dryer in Two Northeast Locations[a] ($/kW, 1980 Dollars)

	Plant A (Two Units)		Plant B (Single Unit)
Vendor[b]	Unit 1	Unit 2	
1	99	91	84
2	103		101
3	117(173)		119(170)
4	(147)		-

[a] Values in parentheses are for a highly urbanized location in northeastern U.S. -- other values are for a generalized location in same region.

[b] Vendors 2, 3, and 4 based their estimates for Plant A on a single FGD system treating the flue gas from both units.

Source: Ref. 23.

First-year operating costs were also estimated for Plant A and B by the same vendors. Some of the principal assumptions in making these estimates area as follows:

- Annual capacity factor: 65%
- Levelized capital charges: 14.7%
- Lime cost: \$61/ton
- Stoichiometric ratio: 1
- Power cost: 10¢/kWh
- Waste-disposal cost: \$10/ton and \$20/ton

The resultant estimates are shown in Table 4.12 where the values are expressed in 1982 values. In all cases the operating costs for Units 1 and 2 of Plant A have been combined to obtain a single estimate. These data show a variation of approximately ±20% for the generalized location. The highly urbanized location would not increase the O & M costs even though the capital costs increased. A doubling of the waste-disposal costs would be expected to add 0.6-0.7 mill/kWh to the annual costs for the conditions assumed in this study.

A final economic assessment of the spray drying/baghouse filtering technology can be found in Ref. 7. This reference also is the source of data presented in Table 2.1 for the base-case technology, and thus a direct comparison of the costs is feasible. In this study it was assumed that the FGD system was to be fitted to a 1000-MW plant consisting of two identical units. Each unit contained five spray dryers, each designed for 25% of the unit gas flow. Thus the philosophy of one redundant module was followed for this design. Estimates were made for plants burning high-sulfur (4%) eastern coal and low-sulfur (0.48%) western coal. The results of these estimates are shown in Table 4.13, where the original values have been escalated to approximate BOY 1980 dollars. In making these estimates it was assumed that lime is used as the sorbent material and that waste disposal is accomplished by putting the dry material in a landfill.

Table 4.12 O & M Cost Estimates for Spray Dryer in the Two Northeastern Locations[a] (mill/kWh, 1982 mills)

	Plant A Waste-Disposal Cost		Plant B Waste-Disposal Cost	
Vendor	@ \$10/ton	@ \$20/ton	@ \$10/ton	@ \$20/ton
1	2.5	3.1	2.6	3.3
2	1.8	2.4	1.9	2.6
3	2(2)	2.6(2.6)	2.3(2.3)	3(3)
4	(1.7)	(2.3)	-	-

[a] Values in parentheses are for the highly urbanized location, with other values applying to the generalized location.

Source: Ref. 23.

Table 4.13 Cost Estimates for Two 500-MW Spray-Drying Systems (BOY 1980 $)

Required Expenditure	Size of Expenditure: High-Sulfur Application	Size of Expenditure: Low-Sulfur Application
Capital investment ($/kW)	66	28
Plant investment ($/kW) (includes general facilities, fees, and contingencies)	107	43
Total capital requirement ($/kW) (includes preproduction costs and AFDC)	137	53
First-year fixed O & M ($/kW-yr)	7	4.8
First-year variable O & M (mill/kWh)	2.9	0.6

Adapted from Ref. 7.

A comparison of the data in Table 4.13 with the data in Table 2.1 shows that, for the high-sulfur application, the total capital requirement for a spray dryer is estimated to be about 77% of that for a limestone wet scrubber. Both O & M requirements are also reduced by a similar amount. It must be emphasized, however, that the estimates for the spray dryer are based on extrapolations from low-sulfur data. These data therefore should be considered uncertain.

The data for the low-sulfur applications show even greater cost differentials for the spray dryer compared to the limestone wet scrubber. The total capital requirement for the spray dryer is estimated to be only 40% of that for the reference technology. Estimates for O & M costs are also significantly lower, with ratios of 64% and 40% for the fixed and variable components, respectively.

4.3.2 Summary of Spray Drying/Baghouse Filtering Economics

As with most developing technologies, a range of cost estimates exists for spray-drying FGD applications. However, some generalizations do appear to be appropriate. For low-sulfur coal applications the total capital requirement can generally be expected to be in the range of $90-120/kW. Capital costs in this range are equal to approximately 85% of the corresponding values for an equivalent limestone wet-scrubbing system. Estimates for high-sulfur coal applications are scarce and have a much greater uncertainty associated with them, but they appear to be 20-25% greater than for low-sulfur coal cases. Operation and maintenance costs for the low-sulfur coal systems generally range from 1-3 mill/kWh. These values are comparable to, but generally slightly less than, the corresponding values for a wet-scrubbing system. The O & M costs for a spray dryer on a high-sulfur-coal application have a great deal of uncertainty due to the lack of experimental data on the required stoichiometric ratio and other important parameters. It can be anticipated, however, that the total O & M costs of a spray-drying system will be greater than those of a wet-scrubbing system for cases where high-sulfur coal is burned.

4.3.3 Potential Applications

The basic spray drying/baghouse filtering technology appears to be applicable to all types of coal, although high-sulfur coal has not been tested as yet. This technology is applicable to both industrial and utility needs.

As of July 1980, nine spray-drying systems totaling 3255 MWe have been contracted for.[20] These systems, all of which will be used in low-sulfur coal applications, are summarized in Table 4.14. With one exception, all of these plants will use lime as the sorbent. As noted previously the Coyote Station system will use soda ash. A fabric-filter system will be used for particulate collection for all applications listed in Table 4.13 except that the Laramie River - Unit 3 will use an ESP.

The Riverside Station plant will be used as a demonstration plant for the spray-drying FGD process with several different coal types. As discussed more fully in Ref. 24, this plant is capable of burning fuels with low, medium, and high sulfur contents. Several different filter fabrics will be tested in this plant, which is scheduled to be in complete service before any other utility has an operative dry-scrubbing system.

4.4 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT

The operating experience gained from the nine utility applications and the four industrial applications will provide information on the applicability of two-stage FGD spray drying/baghouse filtering systems for full-scale operation. This experience could identify areas where design or operating changes may be beneficial. The experience gained at the Riverside Station plant will be especially valuable because it will provide data over a range of coal sulfur contents and for different filter fabrics.

Test efforts such as those described in Ref. 25 have indicated that a partial recycle of the fly ash and reaction products from the baghouse filter can reduce the sorbent requirements for a given SO_2 removal task. Such improved sorbent utilization may be an important factor in making the spray dryer/fabric filter more economically attractive for high-sulfur coal applications.

4.5 INDUSTRIAL-BOILER APPLICATIONS

As noted above, the basic spray-dryer technology is applicable to industrial as well as utility applications. To date, four such systems have been ordered for industrial use and two of these are currently operational. These industrial units are listed in Table 4.15. The Celanese Fibers and Strathmore Paper systems are currently operational and have achieved the guaranteed SO_2 removal efficiency. This information is very valuable in

Table 4.14 Utility Spray-Drying Systems On Order

Plant Name	Utility	Plant Size (MWe)	Sulfur Content (%)	Guaranteed SO_2 Removal (%)	Capital Cost ($/kW)
Coyote Station-1	Otter Trail Power Co.	410	0.78	70	95
Antelope Valley-1	Basin Electric Power Corp.	440	1.22(max.)	78	113
Laramie River-3	Basin Electric Power Corp.	500	0.81(max.)	90	100
Riverside Station-6&7	Northern States Power	65 ea	1 to >4	70 to 90	a
Springerville Station-1&2	Tucson Electric	350 ea	0.69	61	a
Stanton Station	United Power Association	65	Low to intermediate	a	a
Rawhide Station-1	Platte River Power Authority	250	1.3	80	a
Craig Station-3	Colorado-Ute Association	450	0.70	87	100
Holcombe Station-1	Sunflower Electric Corp.	310	low	80	a

[a] Information not available.

Source: Ref. 20.

Table 4.15 Industrial Spray-Drying Systems Operating Or On Order

Plant Owner	Gas Flow (acfm)	Sulfur Content (%)	Guaranteed SO_2 Removal (%)	Capital Cost ($ 10^6)
Celanese Fibers Co.[a]	65,000	1.5 to 2.5	70 to 80	1.25
Strathmore Paper Co.[a]	40,000	2.3 to 3	75	1.40
Univ. of Minnesota (two units)	120,000 ea	0.6 to 0.7	70	3.30
Calgon Co.	57,000	1 to 2	75	1.60

[a] Currently operational spray dryer.

Source: Ref. 20.

the development of spray-dryer technology because it indicates that the required SO_2 removal efficiencies can be achieved with eastern coal containing 2.3% sulfur.[26]

The first three systems noted in Table 4.15 use lime as the sorbent, whereas the Calgon unit is designed for soda ash.

Argonne National Laboratory is currently installing a spray dryer/fabric filter FGD system on the largest of its five boilers. This boiler can produce approximately 170,000 lb/h of steam for heating, hot water, and other services. The installation of this system will allow the boiler, which is currently fueled by gas and oil, to be converted to burn high-sulfur (3.5%) Illinois coal while meeting the State of Illinois atmospheric emission standards. Lime will be used as the sorbent in this system.

Data presented in Ref. 11 show that the capital investment for a spray dryer FGD system is greater than that for the reference sodium throwaway system. For example, a $100 \cdot 10^6$ Btu/h boiler burning low-sulfur coal would require a capital investment of about $1.1 million for a spray dryer as compared to about $0.7 million for the reference system. The capital investment for a spray dryer system increases with boiler size at a much greater rate than does the reference technology. The annualized operating cost for an industrial spray dryer using lime as the sorbent is also estimated to be approximately 50% greater than for the reference technology.

5 AQUEOUS-CARBONATE DRY SCRUBBING

SUMMARY

The aqueous-carbonate process combines the spray-dryer technology with methods of regenerating the sorbent material and producing a marketable end product with the sulfur removed from the flue gas. This FGD technology can therefore greatly reduce the amount of sorbent (Na_2CO_3) required and can also produce revenues that partially offset its own costs. Regenerable and sulfur-recovery techniques make the system complex, however, and this complexity raises both investment costs and operating and maintenance costs. On the other hand, SO_2 removal efficiencies can potentially exceed those that are possible with other advanced FGD technologies. A 100-MW test facility is currently under construction.

5.1 TECHNICAL ASPECTS

5.1.1 Process Description

Another FGD system using spray-dryer technology in combination with basic scrubbing is the aqueous-carbonate process developed by Atomics International. A simplified flow diagram for this process appears as Fig. 5.1. Unlike the processes discussed thus far, this system is intended to regenerate the sorbent material and to recover sulfur for resale.

After the flue gas leaves the air heater, it passes through a cyclone separator, where the ash content is reduced to 1 g/scf or less. The gas then goes to a spray dryer, where it comes in contact with atomized droplets of Na_2CO_3 solution. The sodium carbonate reacts with the SO_2 in the flue gas to yield sodium sulfite and sodium sulfate. The heat of the flue gas evaporates the water; the solid matter (including Na_2SO_3, Na_2SO_4, Na_2CO_3, and fly ash) is carried in the flue gas to another cyclone and an ESP; and those devices remove the solid matter before the gas is discharged through the stack. Although Fig. 5.1 shows that the gas is reheated prior to discharge, this may not be needed with proper design of the spray dryer.

The dry products from the second cyclone are carried to a reducer, where they are heated to about 1800°F. At this temperature the carbon in the coal reacts with the sodium salts and reduces them to sodium sulfide (Na_2S). The melt is quenched and dissolved in water, from which the insoluble impurities are filtered out. The sodium-sulfide solution is pumped to a carbonation area, where CO_2 off-gas from the reducer regenerates an Na_2CO_3 solution and produces H_2S which can then be used to yield sulfur.

Fig. 5.2 (from Ref. 2) shows the material balance for this process. The specific application shown in this figure is an 800-MW plant burning 3.5%-sulfur Illinois coal. Other assumptions made in developing the material balance are noted under the figure.

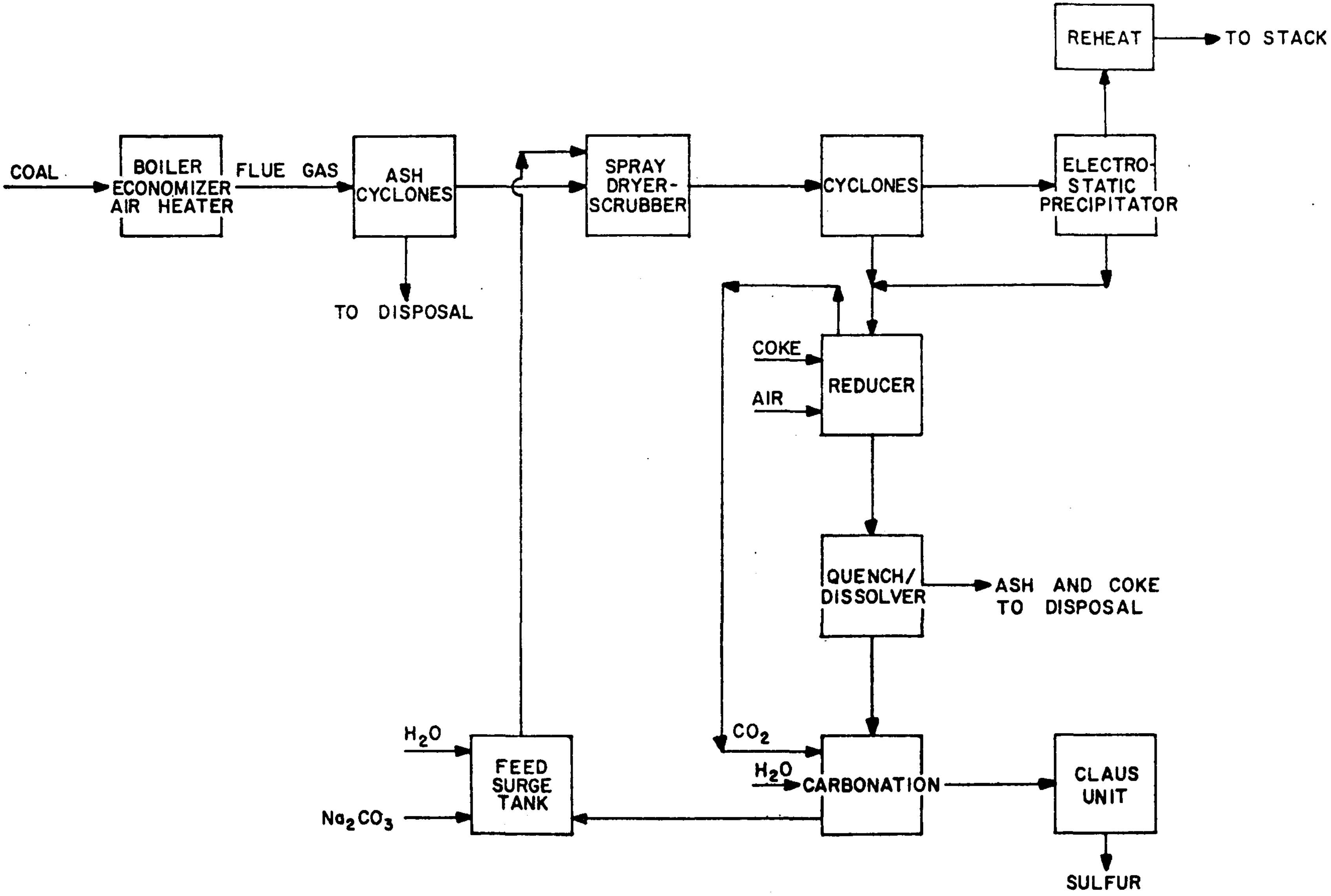
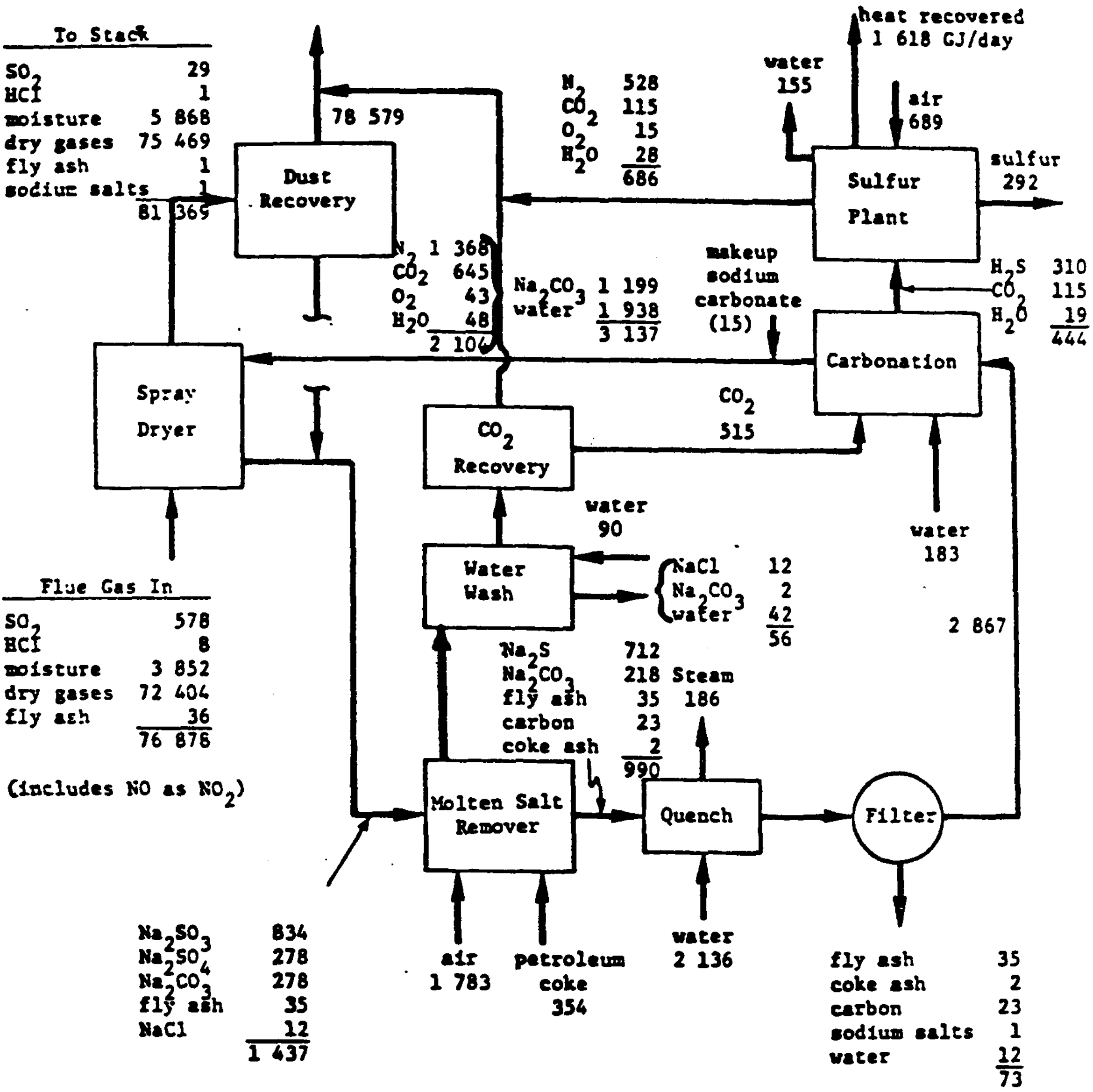


Fig. 5.1 Flow Diagram of Aqueous-Carbonate Dry-Scrubbing Process (Source: Ref. 4)



Values Stand for Metric Tons per Day

Fig. 5.2 Material Balance of an Aqueous-Carbonate Scrubber (Based on 95% Removal of SO_x; 80% Utilization of Alkali; and Oxidation to 20% Sulfate, 60% Sulfite, and 20% Carbonate) (Source: Ref. 2)

5.1.2 Process Chemistry

The chemical reactions involved in removing the SO_2 from the flue gas are as follows:

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2 \qquad (5.1)$$

and

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4 \qquad (5.2)$$

5.1.3 Process Performance

SO_2 Removal Efficiency. Reference 2 notes that the SO_2 removal efficiency for the aqueous-sodium-carbonate process can be 90-95% when a stoichiometric ratio of 1.2 is used.

System Reliability. No information is available at this time. However, the addition of the regeneration and sulfur-recovery sections can be assumed to have an adverse effect on system reliability as compared to a more basic spray-dryer system.

Sulfur Recovery. Recovery of elemental sulfur is an integral part of this process and was discussed in Sec. 5.1.1 above.

Effect of Coal Type on Performance. No information is available at this time.

Effects of Plant Size and Operating Mode on Performance. No information is available at this time.

5.1.4 Water, Chemical, and Power Requirements

The water and chemical requirements for this process are noted in Fig. 5.2. Because this process regenerates the sorbent, only a small amount of Na_2CO_3 is required as makeup. It is estimated that the total energy required for this process is equal to about 5.9% of the power plant fuel.[2] About 80% of this energy is in the form of fuel used in the regeneration process. The regeneration cycle also has a high water requirement for cooling, washing, etc.

5.2 ENVIRONMENTAL IMPLICATIONS

5.2.1 Effluents

The effluents for this process are noted in Fig. 5.2.

5.2.2 Waste Disposal

Because the process incorporates both sorbent regeneration and sulfur recovery, the waste-disposal requirements are significantly reduced. As indicated in Figs. 5.1 and 5.2, the principal product to be disposed of is fly ash.

5.3 ECONOMIC AND MARKET FACTORS

5.3.1 Operating and Capital Costs

Cost estimates for an 800-MW plant burning 3.5%-sulfur Illinois coal are presented in Ref. 2. These estimates have been adjusted to approximate BOY 1980 dollars and are summarized in Table 5.1. The high capital investment is indicative of the complexity of the sorbent-regeneration and sulfur-production processes. Estimates for the reference FGD technology (limestone wet-scrubbing) were also given in Ref. 2 and are discussed in Sec. 2.3.1. The capital investment for the aqueous-carbonate system is 46% greater than that of the reference technology. Fixed O & M costs for the aqueous-carbonate process are more than double those for the limestone system. Annual O & M costs are estimated to be 70% more than those for the reference technology even after a credit for selling the recovered sulfur is applied.

Table 5.1 Cost Estimates for 800-MW Aqueous-Carbonate System (BOY 1980 $)

Required Expenditure	Size
FGD capital investment ($/kW)	182
Investment for replacement power ($/kW)	33
Total capital investment ($/kW)	215
Fixed O & M ($/kW-yr)	16
Variable O & M (mill/kWh)	2.7
Total O & M (mill/kWh)[a]	4.6

[a] Assuming a 70% capacity factor and a credit of $45/ton for sulfur recovered and sold.

Adapted from Ref. 2.

Estimates are provided in Ref. 8 for a 500-MWe plant burning 3.5% sulfur coal. A 90% SO_2 removal efficiency was required for this plant. When adjusted to approximate BOY 1980 values, the total plant investment was estimated at $225/kW. A total capital requirement (including allowance for funds during construction, preproduction costs, etc.) of $274/kW was estimated. It must be noted that in a review of this reference, Atomics International commented that these values are too high and that values of $150/kW and $173/kW are more appropriate.

A fixed annual operating cost of $17.90/kW-yr was estimated for the 500-MW plant. Variable O & M costs of 2.51 mill/kWh were also estimated. These figures do not include waste-disposal costs. At a 70% capacity factor, the total O & M costs estimated for this plant are therefore 5.44 mill/kWh.

In summary, the cost estimates for this FGD process show a wide range. This efffect is at least partially due to the relative newness of the process. Total capital cost estimates of $173-274/kW have been made. Operation and maintenance costs of 4.6-5.4 mill/kWh have also been made.

5.3.2 Potential Applications

Although no specific technical information has been found on this matter, it appears that the aqueous-carbonate process should be applicable to all types of coal. In theory, it should be applicable to both utility and industrial users, but specific economic factors would govern.

5.3.3 Market Potential

No information is available at this time.

5.4 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT

Advances in the aqueous-carbonate FGD technology can be expected to occur as a result of test programs such as the one described below. Information on system reliability and operation and maintenance requirements can be expected. As experience is gained, more efficient plant designs and material utilization can be expected.

A test program is currently underway at the Huntley generating station of Niagara Mohawk Power Corp. in New York. This program, which is sponsored in part by the U.S. Environmental Protection Agency, covers the design, construction, and operation of a 100-MW aqueous-carbonate FGD system. The boiler burns coal containing 2.5% sulfur. An SO_2 removal efficiency of greater than 90% is a design objective. The gas generated in a Claus plant will also be scrubbed in this system. This facility is currently under construction.

5.5 INDUSTRIAL-BOILER APPLICATIONS

Conceptually, this process is applicable to industrial boilers as well as utilities. However, the high capital investment and overall system complexity make it unlikely that aqueous-carbonate scrubbing will make any significant penetration into the industrial market.

6 COPPER-OXIDE DRY SCRUBBING

SUMMARY

The dry copper-oxide process offers the potential of significant NO_x removal as well as very good SO_2 removal. It is a regenerable process that produces a marketable end product. Capital expenditures for this process are expected to be very high. The energy required to regenerate the reactor beds so that they will accept additional SO_2 is estimated to be a significant fraction of the powerplant output. This FGD process is currently in the pilot-scale testing phase.

6.1 TECHNICAL ASPECTS

6.1.1 Process Description

The Universal Oil Products Co. (UOP) markets a dry FGD system that utilizes the Shell copper-oxide process. In this process, a simplified flow diagram of which is shown in Fig. 6.1, the SO_2-laden flue gas flows over the reactor bed, rather than through it as in most other processes. At the same time it removes SO_2, this process removes NO_x by injecting ammonia into the gas stream. The ammonia reacts with NO to form nitrogen and water vapor. Because of this additional feature, the Shell copper-oxide process has been examined intensively in foreign countries, particularly Japan.

Flue gas leaves the boiler at about 750°F. Ammonia is injected into the gas stream, which then flows into one of the reactors. The system will have multiple reactors in parallel. Flue gas enters a reactor that is in the acceptance stage, and flows over copper acceptor material resting on a special support of Al_2O_3. As the flue gas moves through the reactor, the copper is oxided to CuO, which in turn removes SO_2 from the flue gas.

The flue gas then leaves the reactor, flows through the air preheater and such particulate-removal apparatus as an ESP, and is discharged through the stack.

When acceptor material in one of the reactors becomes saturated with SO_2, to the point that its removal efficiency is limited, the flue gas is diverted to another reactor and the original reactor is regenerated. Regeneration is accomplished by passing a steam-diluted, hydrogen-containing gas over the reactor. This regeneration gas can be produced by several processes, including steam-naptha reforming and coal gasification. Regeneration is done at the same temperature as that used in the acceptance phase (about 750°F) so that there is no need to heat or cool a reactor as it cycles from one mode to the other.

The regeneration off-gas contains SO_2, water vapor, and traces of unreacted reducing gas. To recycle this off-gas for regeneration purposes, the bulk of the water vapor must be condensed. To accomplish this, the

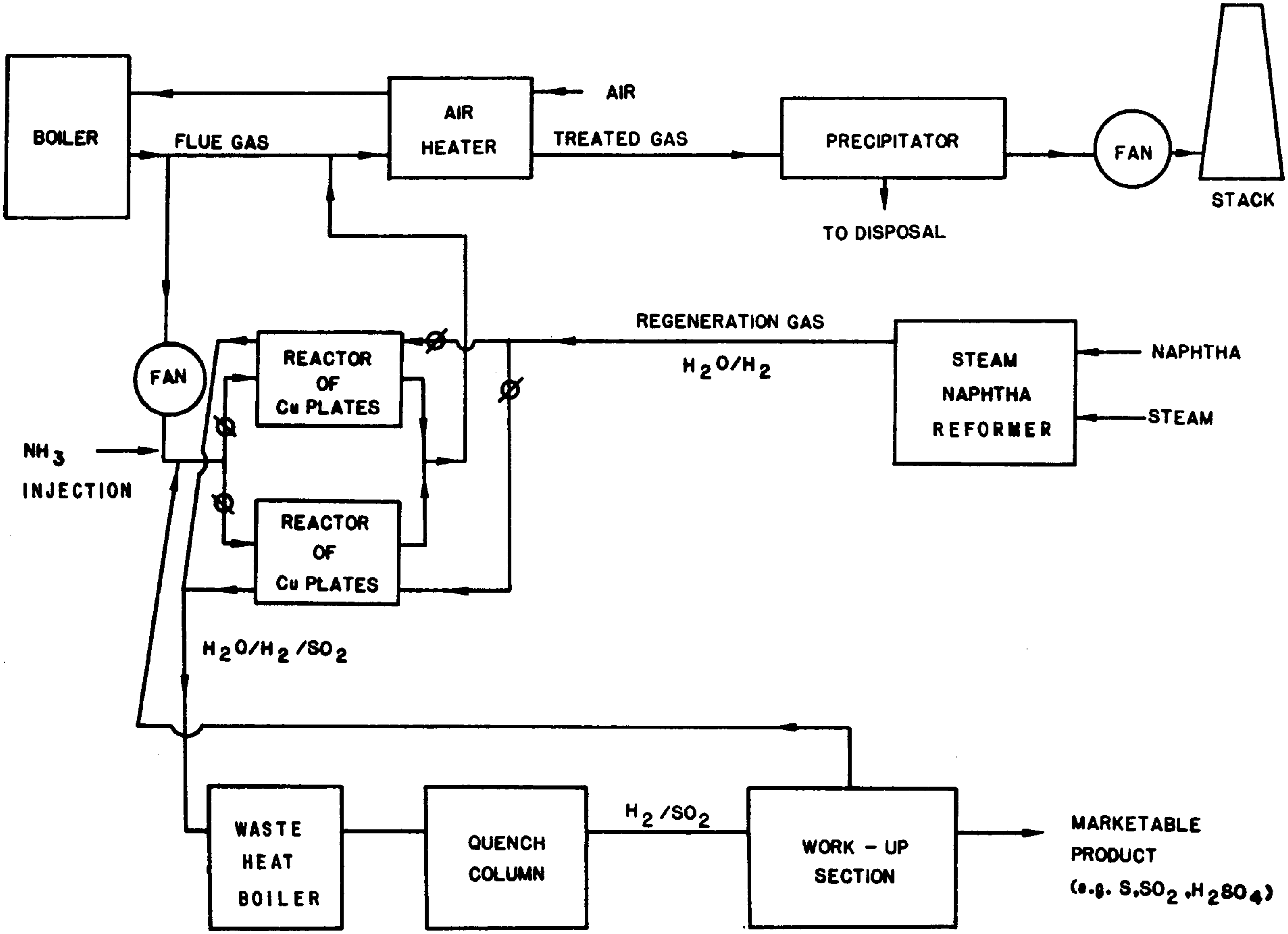


Fig. 6.1 Flow Diagram of Copper-Oxide Dry-Scrubbing FGD Process

off-gas is passed through a waste-heat boiler to recover its sensible heat and then into a direct-contact cooler for heat removal below the dew-point temperature. Upon leaving this equipment, the gas, now SO_2-concentrated, goes to a workup section where any of the following may take place: (1) production of elemental sulfur, (2) production of liquid SO_2, or (3) production of H_2SO_4. The choice between these processes depends on the local costs and product demand.

Fig. 6.2 (taken from Ref. 2) shows the material balances for the copper-oxide FGD system as applied to an 800-MW plant burning Illinois coal with 3.5% sulfur.

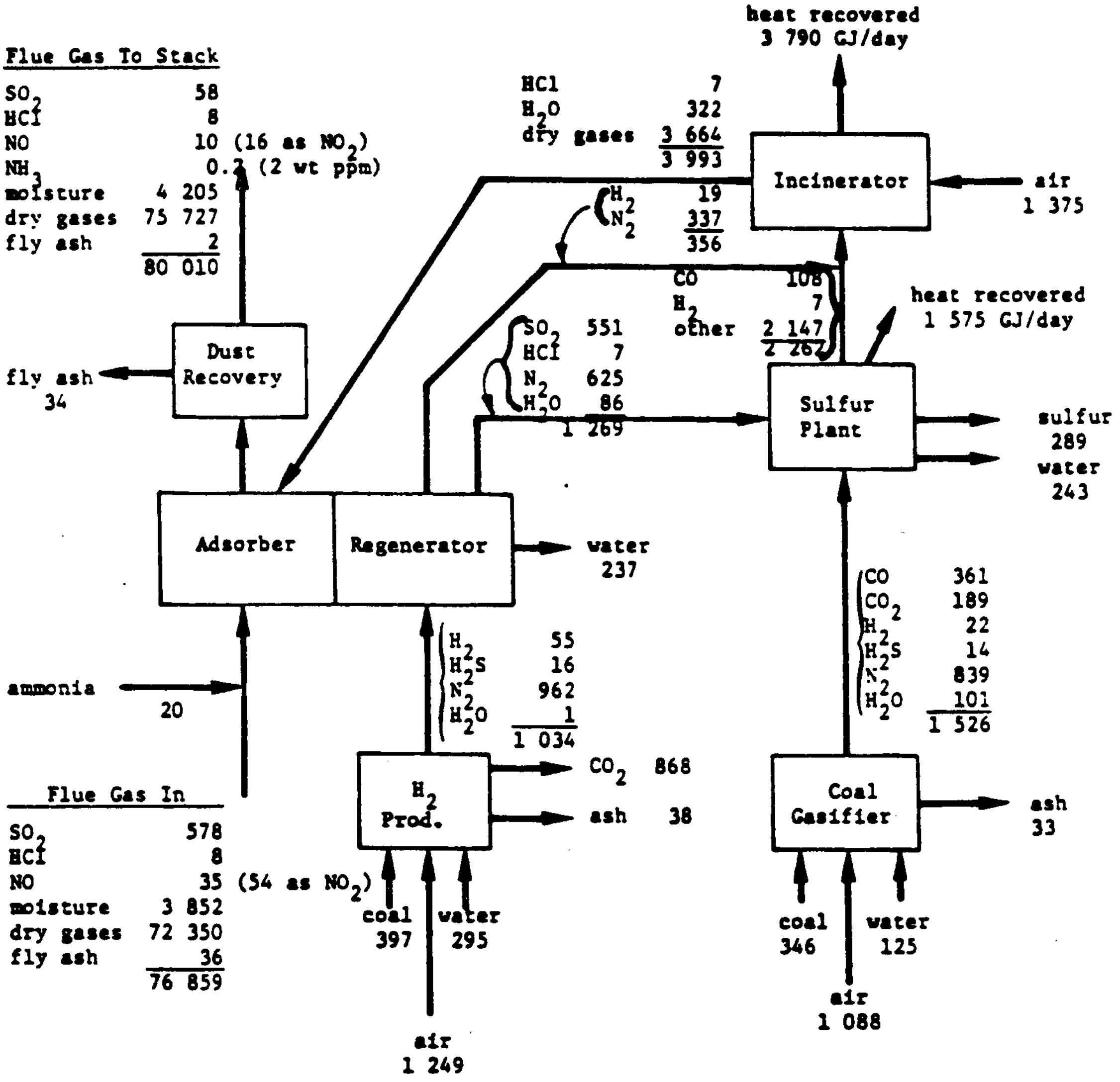
6.1.1.1 Test Programs

The following summary of test efforts conducted to date is taken primarily from Ref. 27.

Laboratory bench-scale testing on this basic process was begun by Shell in the early 1960s. A unit with a capacity of 0.2-0.3 MW was built in 1967 at Pernis, near Rotterdam, in the Netherlands. High-sulfur fuel oil was burned at this plant. It operated for four years and underwent more than 20,000 cycles. From the Pernis operation, it was shown that a diluted H_2-containing gas is preferred for regeneration. This facility also indicated that the physical and chemical stability of the acceptor was excellent and a lifetime in excess of 8000 cycles could be expected. This facility also provided data on the corrosion rates for several metals under conditions expected during FGD applications. An important factor in the operating parameters (and therefore the subsequent cost) of a copper-oxide dry-scrubbing system is the pressure drop across the reactor. Experience at Pernis showed the pressure drop to be low and constant over the reactor life.

An equivalent 40-MW system was installed at Showa Yokkaichi Sekiyu (SYS) in Japan in August 1973. This unit was oil-fired and produced flue gas containing 2500 ppm SO_2. About 90% SO_2 removal was realized at this unit. The NH_3 injection equipment was permanently installed in mid-1975, and the unit has functioned with simultaneous NO_x-SO_2 removal. Valuable experience on the operating characteristics of a larger system was gained at SYS. For example, it was learned that individual cells could be handled and loaded into reactors with no difficulty. It was also realized that an engineering scaleup in size consisted principally in placing standard unit cells in parallel. Special valves to isolate the reactors during regeneration worked satisfactorily at SYS, as did a bypass system designed to make FGD and boiler operations independent of each other. The entire unit could be operated by an automatic-sequence controller, thus requiring very little operator attention after the startup periods. The actual performance of the SYS system agreed with computer projections, thus indicating that the process was well understood and that computer programs could be used for reactor design and optimization.

A 0.6-MW pilot plant of Tampa Electric Co. (TECO) began operation in 1974. This plant burns coal containing 3.5% sulfur. The flue gas for this unit could be taken either upstream or downstream of the ESP. The TECO plant showed that the acceptor was chemically and physically stable after 13,000



Values Stand for Metric Tons per Day

Fig. 6.2 Material Balance of a Copper-Oxide Scrubber (Based on 90% Removal of SO_x and 70% Removal of NO_x when NH_3 is Added at a Rate of 1 mol/mol NO_x, and Oxidation to 100% Sulfate) (Source: Ref. 2)

cycles (equivalent to a three-year life) with 90% SO_2 removal. The bed in this system was 4 m long. A pressure drop of 5-6 in. in H_2O existed across the reactor. The unit performed well with a particulate concentration of 10 g/scf. With this concentration in situ cleaning of reactor internals was required and developed.

6.1.2 Process Chemistry

The basic chemical reaction which removes SO_2 from the flue gas during the acceptance phase is:

$SO_2 + 1/2\ O_2 + CuO \rightarrow CuSO_4$ (6.1)

The copper sulfate produced in this reaction then serves as a catalyst in the removal of NO via the reaction:

$6\ NO + 4NH_3 \rightarrow 5N_2 + 6\ H_2O$ (6.2)

During the regeneration phase the basic reactions are:

$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O$ (6.3)

$CuO + H_2 \rightarrow Cu + H_2O$ (6.4)

6.1.3 Process Performance

SO_2 Removal Efficiency. Test data indicate that about 90% SO_2 removal efficiencies may be possible with this process. UOP states that 90% NO_x removal can also be achieved.

System Reliability. Although no specific data are available for large-scale applications, the experience gained with small units indicates that the system reliability may be acceptable. Furthermore, it was realized that the method of scaling up the technology consists principally of placing standardized unit cells in parallel.

Sulfur Recovery. As indicated in Figs. 6.1 and 6.2 and discussed in Sec. 6.1.1 above, some form of sulfur recovery is an integral part of this process.

Effect of Coal Type on Performance. It is believed that this process can be used for coals of different sulfur contents by varying the number of unit cells and the ratio of the cells in acceptance to cells in regeneration at any given time.

Effects of Plant Size and Operating Mode on Performance. As noted earlier, the number of unit cells would be increased for a larger plant, and thus plant size would not be expected to be a significant factor in the performance of this FGD system. A full turndown capability is claimed by UOP.

6.1.4 Water, Chemical, and Power Requirements

The material balance for this process was shown in Fig. 6.2. Reference 2 estimates that the power requirement would be about 10.6% of the total power-plant fuel. Almost 95% of this energy is in the form of fuel used in the regeneration and sulfur-recovery sections.

6.2 ENVIRONMENTAL IMPLICATIONS

6.2.1 Effluents

The effluents generated in this process are summarized in Fig. 6.2. It can be seen that there are no significant waste-disposal requirements for this system other than those from the particulate removal and those associated with producing a marketable product from the SO_2. An environmental concern could exist if there were to be excessive leakage of NH_3 into the atmosphere, but UOP states that the NH_3 concentration in the treated gas averages 1 ppm.

6.2.2 Waste Disposal

The waste-disposal requirements for this process consist principally of those associated with fly ash.

6.3 ECONOMIC AND MARKET FACTORS

6.3.1 Operating and Capital Costs

Reference 27 presents estimates for a 500-MW plant at a midwestern U.S. location. In making these estimates, it was assumed by UOP that the system had eight separate reactors, with six in the acceptance stage and two in the regeneration stage at all times. Each reactor contains 54,000 kg of acceptor material and has a bed length of 6 m. Flue gas at 750°F enters the reactors at 1,657,000 m^3/h. The SO_2 content of the uncleaned flue gas is 2580 ppm, while the NO_x content is 634. Removal efficiencies of 90.5% for SO_2 and 90% for NO_x were assumed. The modified Claus method of elemental sulfur production was assumed as was a steam-naphtha regeneration system. The capital investment for this system was estimated at $159/kW (BOY 1980 dollars).

In estimating the annual costs, it was assumed that the ammonia usage would be 21 ton/day. Naphtha usage was taken to be 57.23 bbl/h. A net steam requirement of 187 ton/d was based on 867 ton/d for FGD and NH_3 injection and a credit of 680 ton/d for the workup, modified Claus, and steam-naphtha units. The total electricity requirement of 7712 kW is equivalent to 1.5% of the total output of the power plant. A heat credit of $131 \cdot 10^6$ Btu/h from fan compression, reaction heat from both acceptance and regeneration stages, and dew-point suppression was claimed by UOP.

Fixed O & M costs are estimated to be approximately $3.6/kW-yr. The variable O & M value is 2.6 mill/kWh. Considering the complexity of this system, these estimates appear low.

Cost estimates for this process are also given in Ref. 2, where the basis for the estimates includes: an 800-MW plant burning 3.5%-sulfur coal; 90% SO_2 removal and 70% NO_x removal; replacement electricity investment of $872/kW; fuel cost of 11.4 mill/kWh, a by-product credit of $45/ton of elemental sulfur. The total investment for the copper-oxide FGD system was estimated to be $181/kW, with an additional investment for replacement electricity of $16/kW. The total investment is thus $197/kW. Thus, the total capital investment indicated in Ref. 2 is about 24% greater than that provided in Ref. 27. The numbers presented in Ref. 2 also indicate that the electricity requirement per installed MW of FGD capacity is about 20% greater than was used in making the estimates given in Ref. 27. Although a complete set of input used in making cost estimates is not given in Ref. 27, it is possible that the "midwestern U.S. location" plant may be using coal of 3-3.1% sulfur content as opposed to the 3.5% sulfur used in Ref. 2. Such a difference in sulfur content could account for the different cost estimates.

The fixed O & M costs for the copper-oxide dry scrubbing process are estimated at $11.3/kW-yr and the variable O & M at 1.5 mill/kWh. The latter figure includes a credit for sulfur production equal to $45/ton.

Comparisons of the above estimates with comparable ones for the reference technology show that the capital investment and the fixed O & M costs are 30-50% greater for the copper-oxide process. The variable costs appear to be reasonably similar after the credit for sulfur production is taken. In general, however, conclusions regarding the costs associated with this process must be rather tentative because of the scatter in the limited available data.

6.3.2 Potential Applications

As noted earlier, copper-oxide scrubbing can be applied to all ranges of sulfur content by adjusting the number of parallel unit cells. The process is applicable in theory to both utility and industrial requirements, with economics again being a determining factor.

6.3.3 Market Potential

No information is available at this time.

6.4 INDUSTRIAL-BOILER APPLICATIONS

In theory this process of FGD can be applied to industrial boilers as well as to utilities. However, as is common to all regenerative systems, the complexity and high costs of the copper-oxide process will likely limit its industrial application.

7 BERGBAU-FORSCHUNG DRY SCRUBBING

SUMMARY

The Bergbau-Forschung activated-char process has been tested and developed to combine high levels of SO_2 and NO_x removal with sorbent regeneration and sulfur recovery. Although the complexity of the system might be expected to result in high capital investments, the economic estimates to date show a wide range of values. Removal efficiencies for SO_2 are expected to be as high as 95%, even for high-sulfur coals. This process is currently being tested on units ranging from 20 to 50 MW. Systems for still larger units have been designed in which the principal parameters of the adsorption section were kept constant and modular construction was utilized in the desorption section.

7.1 TECHNICAL PERFORMANCE

7.1.1 Process Description

The Bergbau-Forschung activated-char process is a technique of accomplishing both SO_2 and NO_x removal using dry-scrubbing technology. This process consists of three basic sections (see Fig. 7.1 for a simplified flow diagram) which accomplish adsorption, regeneration (desorption), and off-gas treatment.

The flue gas exits the air preheater, passes through a highly efficient ESP, and then enters the adsorber. The adsorber is a cross-stream or a counter-flow moving-bed reactor containing activated char. The char, which is made from coal, is in pellet form -- each pellet is 3/8 in. diameter and 1/2 in. long. The char removes the SO_2 from the flue gas as well as the ash that was not collected in the ESP. The cleaned flue gas is then released in a stack without requiring reheat.

A more detailed description of the adsorption and regeneration (or desorption) sections is shown in Fig. 7.2. This figure, adapted from a similar figure in Ref. 28, represents the throughput for a 300-MWe power plant with a flue gas flow rate of approximately $5.9 \cdot 10^5$ scfm at the inlet to the adsorption section. The particular data in this figure are for a unit using the Deutsche Babcock AG design but are indicative of other designs as well.

Upon exposure to the flue gas, the char becomes laden with SO_2, the dust which has passed through the ESP, and a small amount of fines produced by wear on the char. These dust particles are screened out before sending the SO_2-laden char to the desorption section. In the desorption section, the saturated char pellets are blended with hot sand so that the pellets are heated to about 1200°F. As indicated in Fig. 7.2, a ratio of about 9-10 lb of sand per pound of char is used in the desorption section. As the char reaches a temperature of about 1200°F, the adsorbed H_2SO_4 in the char

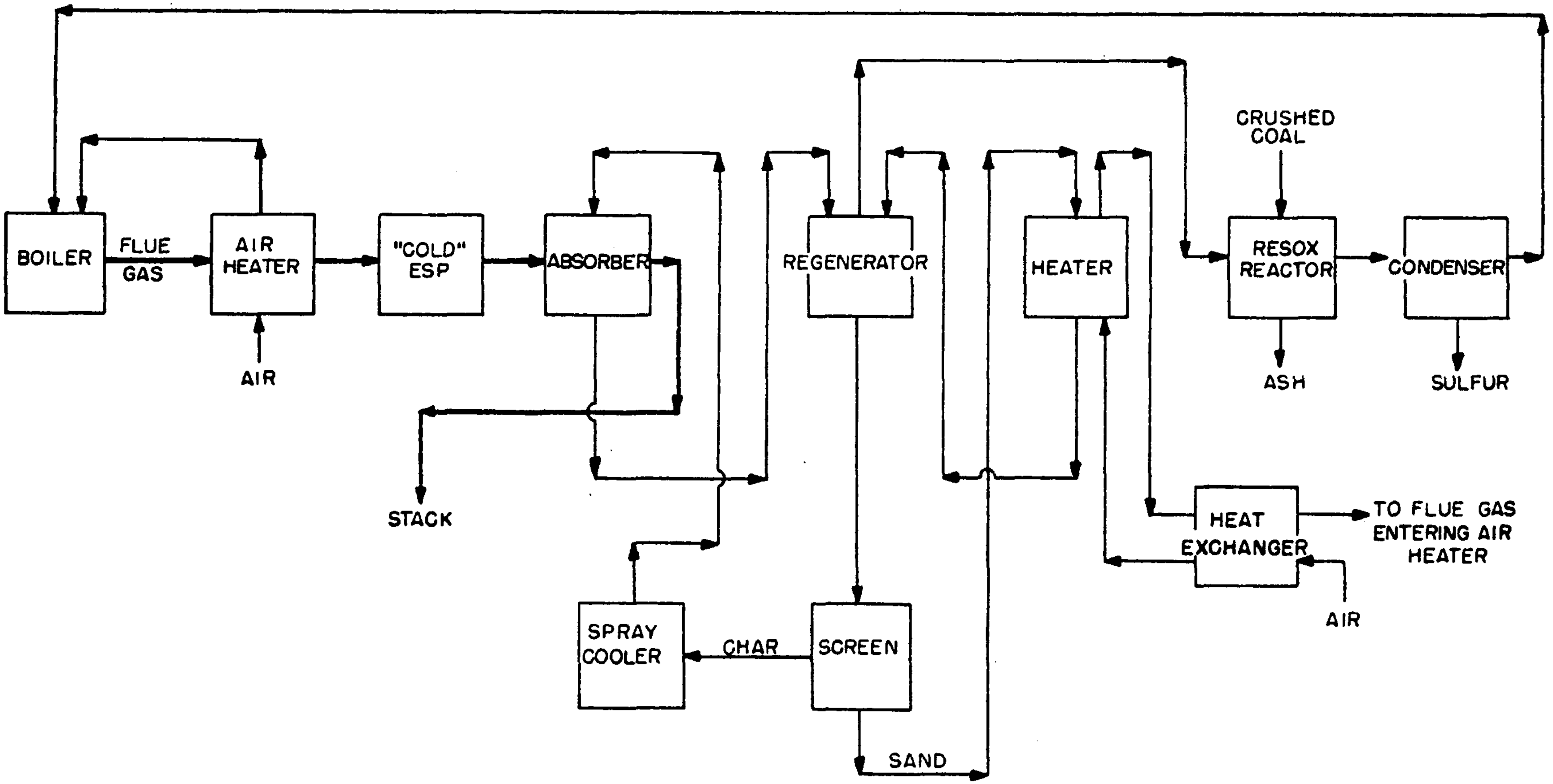


Fig. 7.1 Flow Diagram of Bergbau-Forschung Dry-Scrubbing FGD Process
(Source: Ref. 27)

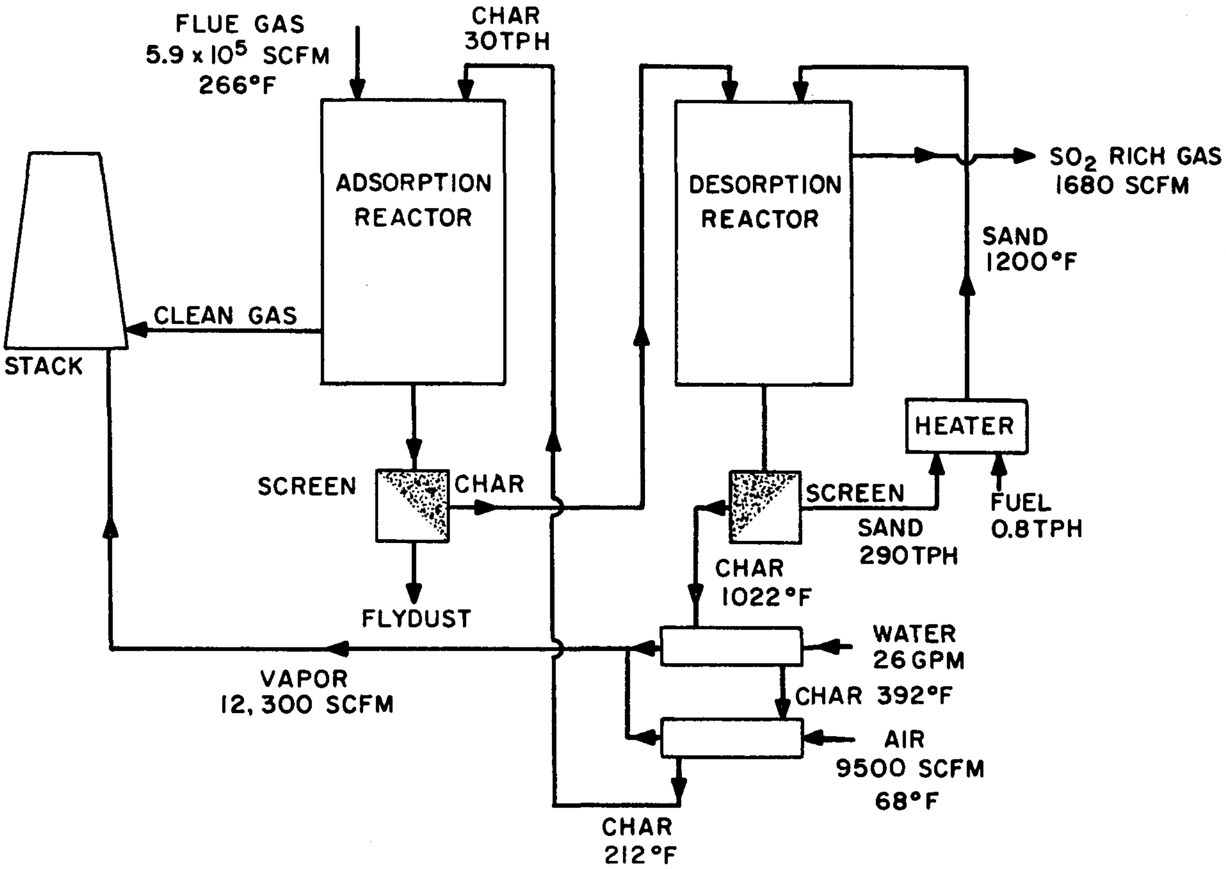


Fig. 7.2 Adsorption and Desorption Sections of Bergbau-Forschung (Activated-Char) Scrubbing Process (Adapted from Ref. 28)

reacts with the carbon to release the SO_2. The adsorbed NO_x is converted to N_2 and released.

A variation to the basic activated-char process is described in Ref. 29. In this process, ammonia is injected into the flue gas in a two-stage adsorber section, and the nitrogen oxides are reduced catalytically to N_2 and H_2O. This process has been demonstrated to remove 80-85% of the NO_x while at the same time enhancing the SO_2 removal efficiency to greater than 95%.

After leaving the desorption section, the regenerated char is cooled to 212°F in two stages. The first stage uses water cooling and reduces the char temperature to 392°F. The second stage utilizes air cooling to lower the char temperature to the desired temperature.

The sand is separated from the char in the screens below the desorption section. This sand is conveyed to a heater where its temperature is raised to in excess of 1200°F for return to the desorption section. The sand is inert to the reactions in the regenerator and is used only as a heat-transfer medium. The heat in the exhaust gases produced in heating the sand is partially recovered in a heat exchanger for the air intake to the desorption heater.

The effluent gas from the desorption section contains 25-30% SO_2 by volume. This gas is passed through a dust separator and then to the SO_2-processing section. This section could be a modified Claus plant for the production of elemental sulfur, a liquefaction plant producing SO_2, or a sulfuric-acid plant. The selection of which processing section to use is dependent upon production costs and market conditions.

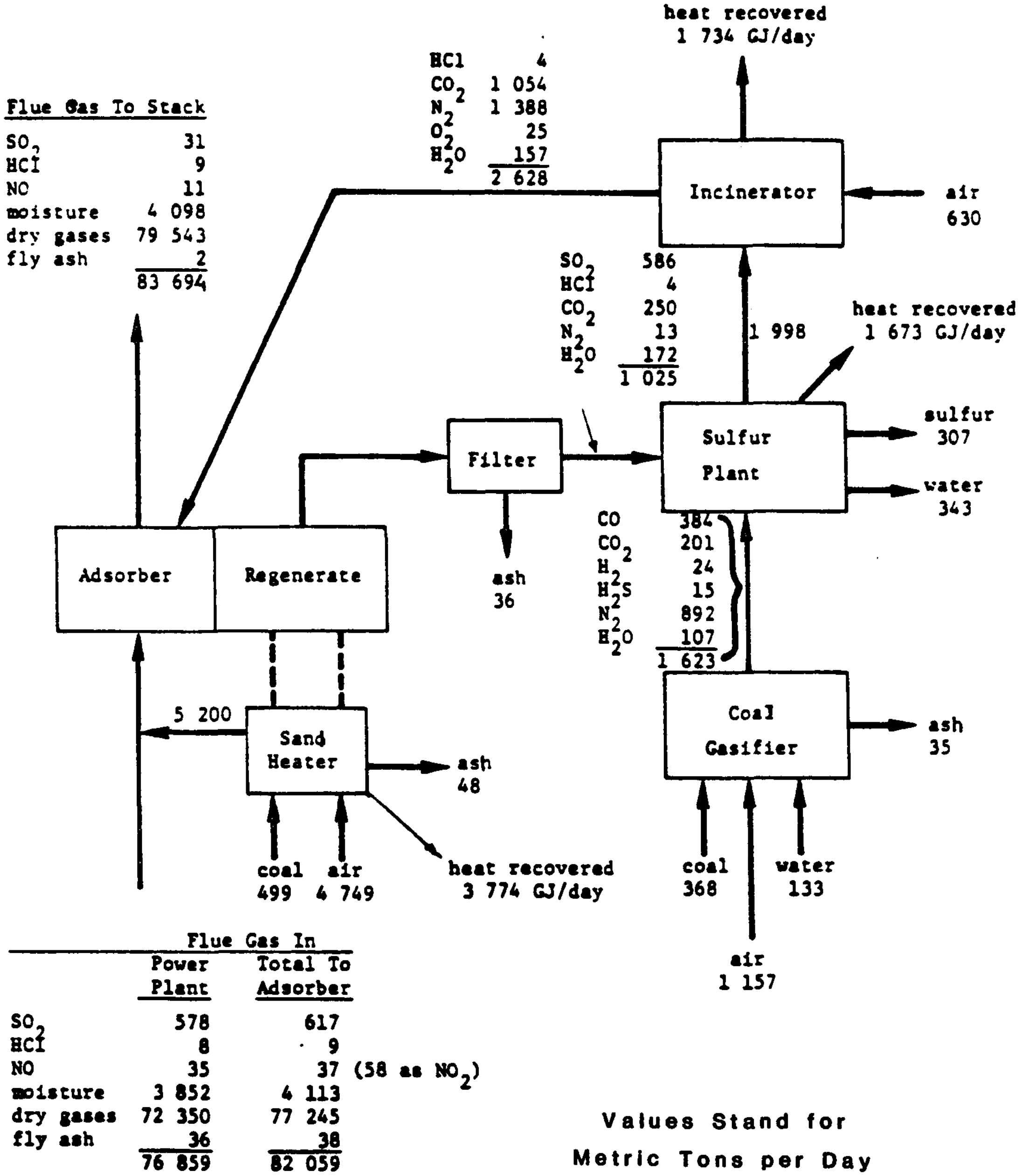
Data on the adsorption and desorption sections of 300-MWe power plants are shown in Fig. 7.2. Figure 7.3 shows the material balance when this process is applied to an 800-MW boiler burning Illinois coal with 3.5% sulfur. In formulating the latter balance, it was assumed that the FGD system would remove 95% of the SO_2 and 70% of the NO_x in the flue gas.

7.1.1.1 Test Programs

Work was initiated on the process in Germany in about 1965. Experience on bench-scale tests and semipilot-plant tests led up to a 105,000 scf/h pilot plant at Welheim, Federal Republic of Germany. This plant, which used flue gas from a coal-fired steam generator, operated for over two years. During one continuous period of 6000 hours, SO_2 removal averaged 80-95%. NO_x removal from the small-scale pilot plants in Germany averaged 40-60%. A 150,000 m^3/h (45-MW equivalent) capacity unit is in operation at Kellermann Power Station in Lünen, Federal Republic of Germany. The flue gas source for this FGD system is a slipstream from a 350-MW coal-fired boiler burning coal of about 2% sulfur content. A modified Claus unit is used to process the off-gas from this unit. This demonstration plant operated successfully for more than a year with SO_2 removal efficiencies of 80-95%.

The construction of a 20-MW prototype unit at the Scholz Steam Plant of Gulf Power Co. in Sneads, Fla., was completed in May 1975. This system has a 20-MW adsorber section and 40-MW regeneration and RESOX sections. It was designed to meet the Florida code for SO_2 emissions of 1.2 lb/10^6 Btu from coal of 3% sulfur, 14% ash, and a heating value of 12,400 Btu/lb. It is believed that the emissions code could be met with a sulfur content as high as 5% because of the oversizing of the regeneration and RESOX sections and the inherent flexibility of the system. Other parameters of significance are as follows.[27]

Parameter	Average or Range
Gas flow to adsorber	174,000 lb/h
Gas residence time in adsorber	13 sec
Char flow rate through adsorber	5,300 lb/h
Char dwell time in adsorber	96 h
Char flow rate to regenerator	5,300 lb/h
Sand flow rate to regenerator	180,000 lb/h
Sand-to-char ratio (vol.)	13.9:1
Coal flow to RESOX reactor	100-250 lb/h
Gas flow to RESOX reactor	750-2000 lb/h



Flue Gas To Stack

SO_2	31
HCl	9
NO	11
moisture	4 098
dry gases	79 543
fly ash	2
	83 694

Flue Gas In

	Power Plant	Total To Adsorber
SO_2	578	617
HCl	8	9
NO	35	37 (58 as NO_2)
moisture	3 852	4 113
dry gases	72 350	77 245
fly ash	36	38
	76 859	82 059

Numbers are in metric tons/day.

NOTE: 68 tons/day of activated carbon is consumed during regeneration by reaction with adsorbate.

Fig. 7.3 Material Balance of a Bergbau-Forschung Scrubber (Based on 95% Removal of SO_x, 70% Removal of NO_x, and Oxidation to 100% H_2SO_4) (Source: Ref. 2)

The adsorber at this unit has two char-bed stages. The first stage consists of eight parallel vertical beds (each 6 ft by 6 ft) while the second stage has four smaller beds (4 ft by 4 ft). All twelve beds are about 40 ft tall. The regeneration vessel and the fluidized-bed heater have been used in industrial applications for several years.

Thus far, the testing at the Scholz plant has been limited because of various equipment problems. SO_x removal efficiencies have averaged about 96% for SO_x compositions varying from 900-2150 ppm in the inlet flue gas. The NO_x removal has averaged only 25-30%, but problems have been experienced with the NO_x monitoring equipment.

7.1.2 Process Chemistry

In the adsorber section, SO_2, water vapor, and oxygen are adsorbed on the activated char, and SO_2 is removed via the reaction

$$H_2O + SO_2 + 1/2\ O_2 \rightarrow H_2SO_4 \qquad (7.1)$$

The H_2SO_4 is then held within the pellet pore structure. SO_3 is adsorbed in the same manner. NO_x is adsorbed at the same time, although the exact mechanism is still in question.

In the regenerator, or desorption, section the char is heated to about 1200°F, and the adsorbed H_2SO_4 reacts with the carbon and is removed from the pellets via

$$2H_2SO_4 + C \rightarrow CO_2 + 2H_2O + 2SO_2 \qquad (7.2)$$

The adsorbed NO_x is removed via

$$2NO + C \rightarrow CO_2 + N_2 \qquad (7.3)$$

When the ammonia-injection technique is used, the following mechanism is hypothesized for NO_x removal.[29] The NO_x and/or the NH_3 is adsorbed on the activated char. The NO_x, the NH_3, and oxygen then react to form N_2 and H_2O.

The sulfuric acid produced in the SO_2 removal process reacts with ammonia via the following reactions

$$H_2SO_4 + NH_3 \rightarrow NH_4HSO_4 \qquad (7.4)$$

$$NH_4HSO_4 + NH_3 \rightarrow (NH_4)_2\ SO_4 \qquad (7.5)$$

These salts (ammonium hydrogen sulfate and ammonium sulfate) are deposited on the surface of the activated char.

7.1.3 Process Performance

SO_2 Removal Efficiency. The average SO_2 efficiency for this process is reported in Refs. 27 and 28 to be 80-95%. An average NO_x removal efficiency of 40-60% is also reported.

System Reliability. No specific information is available at this time, although the demonstration plant of Lünen ran successfully for more than 15,000 hours.

Sulfur Recovery. Sulfur recovery is an inherent feature of this process. The exact process utilized is dependent upon expected market conditions.

Effect of Coal Type on Performance. Test and demonstration efforts have been conducted over a range of sulfur content with acceptable removal efficiencies.

Effects of Plant Size and Operating Mode on Performance. Load variations between 20% and 100% can be adjusted for by varying the quantity of activated char in the circuit. These adjustments can be automatically made if desired.

Many of the principal design parameters in the adsorption section are unchanged with a change in overall plant capacity, so that scale problems are not anticipated. For example, the depth of the moving char bed, the residence time of an individual char pellet, and the free-stream velocity of the entering flow gas are all fixed parameters. In the desorption section, modular construction will be utilized so that no difficulties with various plant capacities are expected.

7.1.4 Water, Chemical, and Power Requirements

Figures 7.2 and 7.3 presented the material balances for this process as applied to a 300-MWe plant and to an 800-MWe plant, respectively. Additional data can be found in Ref. 8 for a 500-MWe plant burning 3.5%-sulfur coal. The electricity requirement for the FGD system is estimated to be approximately 1.6% of the net plant output of 500 MW. Expressed in other terms, it represents about 0.6% of the boiler heat input. The amount of anthracite needed in a RESOX unit to produce elemental sulfur is estimated at 0.01 ton/MWh and the quantity of coal required to heat the sand is estimated to be 0.015 ton/MWh. Combined, these two energy requirements are equal to approximately 6% of the boiler heat input.

Due to erosion on the surface of the pellets, the char would be used at a rate of approximately 6 lb/MWh. Sand would be replaced at approximately 1.3 lb/MWh. Treated water would be used at a rate of 7 gal/MWh, and cooling water would be consumed at about 150 gal/MWh.

7.2 ENVIRONMENTAL IMPLICATIONS

7.2.1 Effluents

The effluents resulting from this process are presented in Figs. 7.2 and 7.3. The principal waste product is the ash removed by the ESP and the various screens in the adsorption and desorption sections.

7.2.2 Waste Disposal

Because of the closed-loop design for this process, waste disposal should be no more difficult than for a conventional coal-fired boiler with particulate control.

7.3 ECONOMIC AND MARKET FACTORS

7.3.1 Operating and Capital Costs

Cost estimates for this process extend over a wide range. For example, Ref. 27 reports capital investments for plants of greater than 200-MW capacity would range from $49/kW for 0.9%-sulfur coal to $109/kW for 4.3%-sulfur coal. Foster-Wheeler (a principal investigator of this sytem) states that the most economical utilization for this process is in the 150-200 MW range and that there is little change in the economics for a larger plant. On the other hand, Ref. 2 states that recent economic capital investment data are not available but that it can be speculated that the costs would probably exceed those for the UOP copper-oxide process, which was quoted at $197/kW for 3.5%-sulfur coal. An overall O & M cost for this system was estimated to be 4.4 mill/kWh in Ref. 2.

Another estimate of the costs associated with the activated-char FGD process is found in Ref. 8. The sytem described there was designed to remove 90% of the SO_2 in the flue gas of a 500-MWe plant burning coal of 12,000 Btu/lb heating value and containing 3.5% sulfur. The overall plant heat rate was given as 9000 Btu/kWh. When escalated from the first quarter 1977 period of the original estimate to the beginning of 1980, the total plant investment becomes approximately $66,400,000 ($133/kWe). The inclusion of an allowance for funds during construction, inventory capital, etc., brings the total capital requirement to $171/kWe. The first-year fixed operating costs were estimated at $5.13 million, or $10.3/kW-yr. The variable O & M costs (excluding waste disposal) were estimated at 4.2 mill/kWh.

It is believed that the estimates for Ref. 27 are too low and that a more probable range of total capital investment requirement is approximately $170-210/kWe. Total O & M costs of 4.4-5.9 mill/kWh would be anticipiated with a capacity factor of about 70%.

7.4 INDUSTRIAL-BOILER APPLICATIONS

In theory, the activated-char dry-scrubbing process is applicable for both industrial and utility use. However, the relative mechanical complexity of this process may limit its use to large systems. No cost information for industrial applications has been found.

8 DIRECT LIMESTONE COMBUSTION

Direct combustion of a coal/limestone mixture is being examined as a technique for reduction of SO_2 emissions.[20] The U.S. Environmental Protection Agency (EPA) is currently sponsoring two variations of the technology, both in the early stages of development.

One of the limestone-combustion technologies involves the production and ultimate combustion of coal/limestone pellets. A test program was to be conducted at General Motors' Indianapolis plant, which utilizes a stoker boiler rated at 60,000 lb/h of steam. This technology must be considered to be in the developmental stage because only small boilers have been used for testing purposes.

The other technology being tested under EPA sponsorship deals with the burning of a pulverized coal/limestone mixture in a low-NO_x burner. The lower flame temperatures in such burners apparently enhance the SO_2 removal efficiencies of direct combustion by preventing the glazing of the reagent particles, which occurs at the higher flame temperatures in conventional burners. The glazing of the particles leads to significantly lower reactivity and thus to lower SO_2 removal efficiencies.

The anticipated advantage of limestone combustion is that NO_x, SO_2, and particulate matter can be controlled with a single, low-capital-investment design. Although this technology is currently in its infancy, cost estimates project that its capital investment will compare favorably with baghouse-filtering or spray-drying FGD systems. O & M costs are anticipated to be low, as well, because of the relative simplicity of this system.

If the ongoing research and development for the limestone-combustion technology prove successful, it should be viable in both utility and industrial applications.

9 SODIUM-SULFITE-BASED DOUBLE-ALKALI WET SCRUBBING

SUMMARY

Double-alkali scrubbing is a wet, regenerable SO_2-abatement process combining absorption of SO_2 with aqueous alkali and regeneration of the absorbent with lime or limestone. The double-alkali systems utilize clear sodium-sulfite-based absorption solution. They reduce the problems of plugging, scaling, and erosion. Existing systems remove SO_2 with 90-95% efficiency. While some systems have had mechanical or chemical problems, they have shown themselves reliable; less than 10% of their total operating time has been interrupted with forced outages. This FGD technology has good retrofitting potential based on the small size of its components. The process requires a large land area for disposing of the solid waste it generates. Economically, double-alkali systems appear to be competitive with the wet lime and limestone FGD systems. The process has been commercially applied in the U.S. Three full-scale demonstration systems are operating with coal-fired utility boilers, and several commercial units are in operation with coal- and oil-fired industrial boilers. Further development work is needed to evaluate, characterize, and compare full-size coal-fired demonstration facilities; to test systems using limestone as a regenerant; and to develop methods for upgrading the quality of sludge and strategies to control the multimedia environmental effects of effluents and emissions.

9.1 TECHNICAL ASPECTS

9.1.1 Process Description

The "double-alkali" or "dual-alkali" FGD processes, generally referred to as "indirect" lime/limestone processes, can be divided into two categories: sodium-sulfite-based and nonsodium-sulfite-based. The sodium-sulfite-based double-alkali systems are discussed in this section. The DOWA double-alkali system, one of the nonsodium-sulfite-based systems, is discussed in Sec. 10.

The sodium-sulfite-based double-alkali systems consist of two sections: absorption and regeneration. The SO_2-absorption step utilizes a sodium-based scrubbing liquor. In the regeneration section, the absorbent is regenerated by reacting with lime or limestone. The absorbed SO_2 is converted into a calcium-based sludge. The term "double alkali" refers to the two alkaline raw-material streams required -- lime/limestone and sodium carbonate.

Figure 9.1 shows a simplified process flow diagram for a double-alkali FGD system. Flue gas from the boiler air preheater first enters a particulate scrubber, where particulates and chlorides are removed through contact with a fly-ash slurry. The flue gas next enters an absorber, where the SO_2 is absorbed by a recirculating solution of sodium sulfite. The scrubbed flue gas

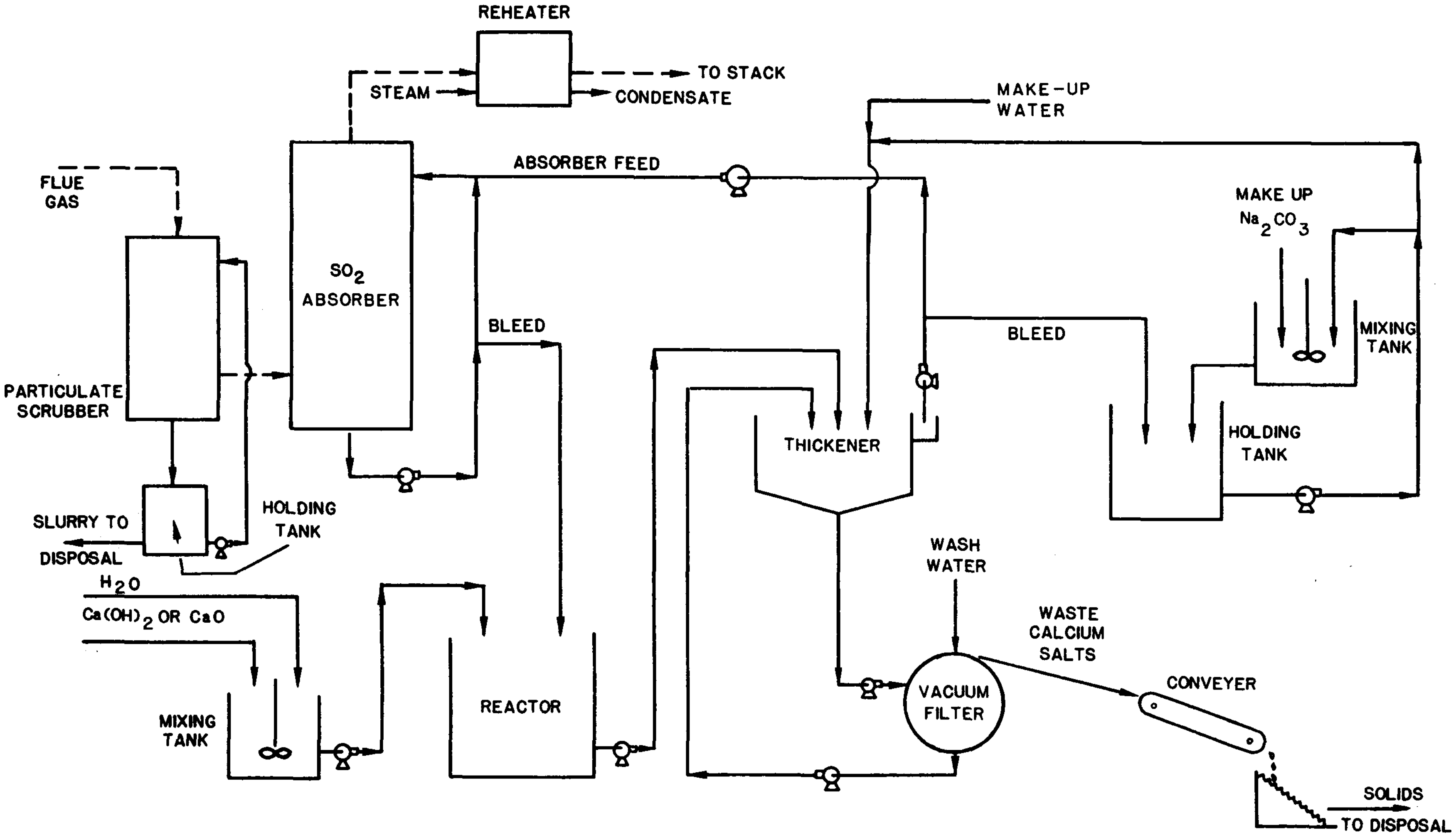


Fig. 9.1 Flow Diagram of Sodium-Sulfite-Based Double-Alkali Wet-Scrubbing FGD Process (Source: Ref. 30)

is heated in a reheater and compressed by an induced-draft fan prior to entering the plant stack for discharge to the atmosphere.

In the absorbent-regeneration system, the absorber-liquor bleed is first sent to a reactor, where the absorbent is regenerated by reacting with lime or limestone. The lime-treated liquor is directed to a thickener, where soda ash is added to reduce the amount of dissolved calcium. The reaction products -- calcium sulfite and sulfate -- precipitate in the thickener, resulting in a slurry containing about 5% solids by weight. Overflow from the thickener is pumped back to the absorber. A slipstream of thickener overflow is sent through the soda-ash mixing tank, where Na_2CO_3 is added to it. The soda ash replaces sodium lost to the system when the filter cake is formed.

The material balance for a full-scale double-alkali demonstration plant (300 MW) at Louisville Gas and Electric Co. is shown in Fig. 9.2. The system was designed by Combustion Equipment Associates, Inc., and A.D. Little, Inc.[30]

9.1.2 Process Chemistry

The chemistry of double-alkali systems is briefly described here. More detailed descriptions are available elsewhere.[31,32]

In the absorption section, absorption of SO_2 in sodium-sulfite solution occurs to produce a bisulfite effluent solution according to the overall reaction

$$Na_2SO_3 + SO_2 + H_2O \rightleftarrows 2NaHSO_3 \tag{9.1}$$

The absorbent feed to the absorber may also contain some sodium hydroxide and/or sodium carbonate. These compounds form sodium sulfite on absorption of SO_2

$$Na_2CO_3 + SO_2 \rightleftarrows Na_2SO_3 + CO_2 \tag{9.2}$$

$$2NaOH + SO_2 \rightleftarrows Na_2SO_3 + H_2O \tag{9.3}$$

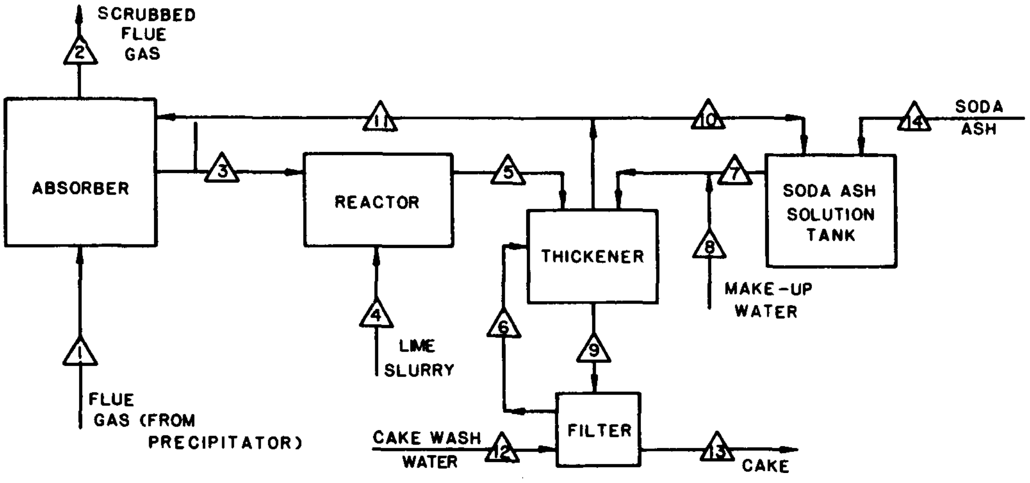
The resulting sulfite is used in further absorption to produce bisulfite. Some oxidation of sulfite to sulfate occurs in the absorber due to reaction of sulfite with oxygen in the flue gas

$$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4 \tag{9.4}$$

The regeneration of sodium sulfite/sulfate effluent solution is achieved by neutralization of the bisulfite using either lime or limestone, producing a precipitate of calcium sulfite

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2H_2O + 3/_2H_2O \tag{9.5}$$

$$2NaHSO_3 + CaCO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2H_2O + CO_2 + 1/2H_2O \tag{9.6}$$



MATERIAL BALANCE : lb/min

STREAM NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14
H_2O	2,632	4,948	27,286	1076	28,466	2,421	404	878	2,109	404	29,656	756	442	
SO_2	390	22												
DRY GAS	53,008	53,008												
PARTICULATES	4	4												
Na_2CO_3							14			0.2	14			14
NaOH					100	7	1		8	1	100		0.5	
Na_2SO_3			240		700	48	10		51	10	694		3	
$NaHSO_3$			912											
Na_2SO_4			2112		2,040	139	28		149	28	2,032		10	
$CaSO_3 \cdot \frac{1}{2} H_2O$					662				662				662	
$CaSO_4 \cdot \frac{1}{2} H_2O$					78				79				79	
$CaCO_3$					12				12				12	
INERTS				35	34				35				35	
NaCl			610		610	42	8		45	8	608		3	
$Ca(OH)_2$				426										
TOTAL	56,034	57,982	31,160	1537	32,702	2,657	465	878	3150	451	33,104	756	1247	14

Fig. 9.2 Material Balance of a Sodium-Sulfite-Based Double-Alkali Scrubber (Source: Ref. 30)

Regeneration can be carried beyond neutralization to produce caustic with lime

$$Na_2SO_3 + Ca(OH)_2 \rightleftarrows 2NaOH + CaSO_3 \qquad (9.7)$$

Depending upon the concentration of sulfite and sulfate and the pH of the solution, the following reaction for sulfate removal also occurs, using either lime or limestone

$$Ca^{++} + SO_4^{=} \rightleftarrows CaSO_4 \qquad (9.8)$$

9.1.3 Process Performance

SO_2 Removal Efficiency. Double-alkali systems have been demonstrated and commercially applied at industrial and utility sites. These systems have demonstrated the ability to perform well over a wide range of boiler, fuel, and flue gas design and operating conditions. With the present state of the art, it is reasonable to expect a long-term average SO_2 removal capability on the order of 90%, a moderate liquid-to-gas ratio, and a high SO_2 concentration at the inlet.

System Reliability. The double-alkali FGD process is somewhat complex and requires more components than the lime/limestone processes. System reliability, however, is expected to be greater than lime or limestone scrubbing processes due mainly to the elimination of scale and plugging problems. The vendors of the process guarantee 90% availability for the first year of operation and, in some cases, for the life of the plant also. More system-reliability information is given in Sec. 9.4 of this report.

Particulate Removal Efficiency. Particulate removal in double-alkali systems can be accommodated by appropriate selection of scrubbers to be used for both the particulate removal and the SO_2-absorption duty. At the minimum, the double-alkali system will not produce net addition of particulate matter to the gas stream.

Dilute Vs. Concentrated Systems. Double-alkali FGD systems are divided into dilute and concentrated systems, depending upon the concentration of active alkali -- i.e., the sum of the concentrations of NaOH, Na_2CO_3, $NaHCO_3$, and $NaHSO_3$. In concentrated systems (active sodium concentration greater than 0.15 m), high sulfite levels prevent the precipitation of sodium sulfate ($CaSO_4$) as gypsum ($CaSO_4 \cdot 2H_2O$). However, $CaSO_4$ is precipitated along with calcium sulfite ($CaSO_3 \cdot 1/2\ H_2O$). In dilute systems, gypsum alone or both gypsum and calcium sulfite will precipitate. Between these two modes of operation, the concentrated systems appear to be more reliable, less complex, and cheaper -- but they are limited to situations where sulfite oxidation rates do not exceed 25-30% of the SO_2 absorbed. The dilute systems can be operated at higher oxidation rates. Discussions in this section are in general pertaining to concentrated systems.

Effect of Fuel Variations. The following factors associated with boiler input fuel can significantly affect the design and performance of a double-alkali FGD system.

- Sulfur Content of the Fuel. This determines to a great extent the potential for sulfite oxidation in the system and consequently affects the selection of design mode (i.e., dilute or concentrated) and operability of the double-alkali system. In general, high-sulfur fuels are expected to have low sulfite oxidation, and concentrated systems are more suited for this situation. Dilute systems are used where low-sulfur fuels (or high sulfite oxidation) are expected.

For fuels containing less than 1% sulfur by weight, a concentrated-mode double-alkali system cannot be operated at the excess air levels that are typical for pulverized-coal-fired boilers without an initial purge of sodium sulfate. However, for fuels containing more than 2% sulfur, and in many cases for coals containing between 1% and 2% sulfur, the operation of the concentrated system is excellent.[33]

- Chloride Content of the Fuel. This is an important factor in the design and operability of a double-alkali system because it affects corrosion potential and SO_2 removal capability of the system. The presence of chlorides in the scrubber liquors provides the potential for stress corrosion which can result in the use of some high-alloy equipment. Dissolved chlorides will also react with active alkalis to form inactive chloride salts and significantly reduce the SO_2 absorption capability of the alkali. These problems can be reduced by adding alkali additives, such as MgO, to the scrubbing liquor. Alternatively, chlorine can be removed ahead of the SO_2 scrubber in a relatively small prescrubber.

- Ash Alkalinity. This factor will influence the design of a double-alkali system in that a highly alkaline ash can significantly decrease the quantity of sorbent required for SO_2 removal. In addition, the ash alkalinity will determine whether or not to remove particulate material ahead of the scrubber.

9.1.4 Water, Chemical, and Energy Requirements

Water. Water is required in the double-alkali process for process uses and noncontact uses. Process water added to the system serves many purposes, including saturation of flue gas, preparation of lime-slurry makeup, waste-product washing, instrument purge, mist-eliminator washing, and pump seals. Process water removed from the system includes evaporation to the flue gas and water in the filter cake. These streams are directly related to the process operation. In addition to these streams, there is evaporation of water to the atmosphere from such open vessels as thickener holding tanks. The net evaporation from open vessels is dependent on the prevailing meteorological conditions in the local area.

The noncontact water streams in the double-alkali systems are the booster-fan cooling water and the seal water for the vacuum pumps. These streams, which are generally segragated from the process streams, can be treated for reuse or for discharge.

Available data appear to indicate that process water requirements for double-alkali systems range from 1 to 1.3 gal/min per megawatt of capacity. Table 9.1 lists the estimated water requirements of a double-alkali system operating on a peak 300-MW capacity coal-fired boiler.[30] The expected operating conditions are 3.8% sulfur coal, 55% solids in the filter cake, and a cake wash ratio of 2. The process-water requirement per unit of capacity was estimated at 1.24 gal/min per MW.

Table 9.1 Estimated Water Flow Rates for a Double-Alkali FGD System[a] (gal/min)

	Flow Rate (Based on 300-MW Unit at Three Load Levels)		
Nature or Source of Flow	20% Load	60% Load	100% Load
Process water inputs			
Lime slurry (free water and	0.9	62.6	104.3
chemically combined water)	2	6	10
Pump-seal water	20.3	21	21.6
Instrument purge water	0.5	0.5	0.5
Filter-cake wash water	24.1	72.2	120.3
Rain water	0.6	0.6	0.6
Total	68.4	162.9	257.3
Noncontact water inputs			
Fan-bearing cooling water	175	175	175
Vacuum-pump-seal water	8	24.2	40.4
Total	183	199.2	215.4
Process water outputs			
Evaporation to the flue gas	55.6	166.9	278.2
Cake (free water and	12	36.1	60.1
chemically combined water)	0.5	1.6	2.6
Evaporation to atmosphere[b]	32	32	32
Total	100.1	236.6	372.9
Noncontact water outputs			
Fan-bearing cooling water	175	175	175
Vacuum-pump-seal water	8	24.2	40.4
Total	183	199.2	215.4
Net water added to system to close process water balance	31.7	73.7	115.6

[a] Basis: 300-MW coal-fired boiler
3.8%-sulfur coal
55% insoluble solids in the filtercake
2 wash ratio

[b] Annual average.

Source: Ref. 30.

Lime/Limestone. Lime or limestone is required in double-alkali processes for absorbent regeneration. A method for specifying the lime/limestone requirement is based on calcium stoichiometry, i.e., moles of calcium added per mole of sulfur removed. A calcium consumption of 0.98-1 mol appears to be a reasonable design target for a concentrated double-alkali system.

A pilot study by Arthur D. Little, Inc., indicated that lime utilization in double-alkali processes easily exceeds 90%; utilization as high as 100% can be achieved.[31] Lime utilization increases with reactor residence time, but it decreases as regeneration is carried beyond neutralization to the generation of free hydroxide, approaching the reaction equilibrium limits.

A double-alkali system operating on a 300-MW boiler with a 60% load factor and firing 3.8%-sulfur coal was estimated to require 59,000 ton/yr to achieve 94% SO_2 removal -- and the lime must contain 70.01% CaO.[30] This requirement is based on the assumption that the consumption of lime in the FGD system will not exceed 1.05 mol CaO/mol SO_2 removed from the flue gas.

Sodium Carbonate. In sodium-sulfite-based double-alkali scrubbing, the sodium makeup requirements must equal the sodium losses in the filter cakes. A method for measuring sodium consumption is in moles of sodium consumed per moles of sulfur removed by the system. It ranges from 0.01-0.05 mol Na_2CO_3/mol SO_2 removed, depending upon the ability to wash the filter cake in the specific applications.[31] A double-alkali system operating on a 300-MW coal-fired boiler was estimated to require 1912 ton/yr of soda ash to remove 94% of the SO_2.[30] This assumes that 0.045 mol Na_2CO_3 are consumed per mole of SO_2 removed.

Energy. Energy required for double-alkali systems includes some combination of electricity, fuel, and steam. The electrical energy is needed for the following five basic operations:

(1) Raw material handling and feed preparation. Areas of electricity utilization include the powering of conveyors, grinders, mixers, and pumps associated with receiving, storing, and preparing makeup lime/limestone and soda ash.

(2) Particulate/chloride removal. The principal electricity use involves the flue-gas pressure drop and equipment power associated with venturi scrubbers to remove both substances.

(3) SO_2 scrubbing. Electrical energy is needed for the operation of pumps, agitators, etc.; to operate the absorber; and to transport materials against pressure drops in the flue gas passing through the scrubber.

(4) Operation of fans. Electrical energy is needed to operate either induced- or forced-draft fans in order to maintain gas flow through the scrubbing systems.

(5) Disposal of solid waste. For onsite disposal, electrical energy is required to pump water and waste material to and from the settling ponds as well as to operate agitators in the feed tanks. For offsite disposal, energy is needed for transportation of solid waste to, and for preparation and reclaimation of, disposal sites.

As for many other FGD processes, heat is needed for double-alkali systems to raise the scrubbed flue-gas temperature prior to discharge to the atmosphere. This is done to prevent the formation of sulfuric acid mist and to provide a sufficient plume buoyancy. Energy for this purpose could be provided by means of steam or combustion fuels, and its quantity can be related to the flue-gas flow rate and temperature rise desired. Scrubbed gas in the U.S. is generally reheated to achieve an exit temperature of 175°F.

Available data appear to indicate that total electrical energy (excluding reheat) required to operate a double-alkali system ranges from 1% to 2% of the electrical generating capacity. Total energy required to operate a double-alkali system including electricity and fuel for reheat, could account for 2.5-4% of total heat input to the boiler.

Table 9.2 summarizes the estimates of electrical energy requirements for a double-alkali system installed at a utility. Electrical energy requirements for the system could derate the power generating system by 7.87 MW (or 1.57% of the generating capacity) of a 500-MW power plant. For the same system, the energy required for raising the temperature at the scrubbed flue gas by 38°F was estimated at $53.2 \cdot 10^6$ Btu/hour. Assuming 88% boiler heat efficiency, the steam rate to the reheater represents 1.34% of total heat input to the boiler. It is important to note that energy requirements of the system depend principally on the flue gas flow rate -- they depend very little on SO_2 content at the inlet or on SO_2 removal rate.[34]

Table 9.2 Estimated Electricity Requirement for a Double-Alkali FGD System[a] (kW)

Area	Requirement
Raw material handling and preparation	100
Particulate/chloride removal	1090
SO_2 scrubbing	860
Fans	5510
Disposal of calcium solids	250
Utilities and services	60
Total	7870 (1.574% power generating capacity)

[a]Basis: 500-MW power plant burning 3.5%-sulfur coal and expected to achieve 90% SO_2 removal.

Source: Ref. 34.

9.2 ENVIRONMENTAL IMPLICATIONS

Gaseous Emissions. As stated earlier, double-alkali systems are able to remove greater than 90% of flue gas SO_2 without producing net addition of particulate matter to the existing gas steam. This FGD process has no capability of NO_x removal.

Aqueous Effluents. The double-alkali systems produce a continuous effluent in the prescrubbing system blowdown and an intermittent effluent in absorbent solution purge. The prescrubbing system blowdown may have a relatively high concentration of chloride and suspended solids and trace amounts of metals. The absorbent purge stream could have high sodium sulfide, sodium sulfate, and nonsulfur calcium salts.[35] These liquid effluents can be treated for discharge or reuse.

Solid Waste. The double-alkali processes produce a solid waste made up mostly of unreacted lime/limestone, calcium sulfite, calcium sulfate, and chemical from fly ash. These substances are also found in lime/limestone FGD solid wastes. In addition to these substances, however, the double-alkali solid waste contains soluble sodium salts such as NaCl and Na_2SO_4 -- these salts cause environmental concerns. The relative amounts of these chemicals depend upon the control system, its design and operating variables, and the type of coal burned.

For a double-alkali FGD system operating with a 500-MW plant burning coal containing 3.5% sulfur and 14% ash, the amount of solid waste produced was estimated at $2.334 \cdot 10^6$ ton/yr.[36] This estimate was based on 80% SO_2 removal, or 1.2 lb $SO_2/10^6$ Btu emission limitation. Half (dry weight) of this solid waste is composed of fly ash. Assuming a solid-waste depth of 30 ft, approximately 300 acres of land will be required for disposal of the waste over a 30-yr period. The land used for solid-waste disposal would be lost to other commercial or agricultural purposes as long as the disposal site is operated.

9.3 ECONOMIC AND MARKET FACTORS

9.3.1 Capital and Operating Costs

Available cost estimates for the double-alkali and several other FGD systems are given in Table 9.3. It appears that the double-alkali system is economically competitive with wet lime and limestone systems, and less expensive than several other FGD systems such as MgO, Wellman-Lord, and Citrate. These conclusions, of course, are subject to change if the basic assumptions used to estimate cost require significant revision.

A detailed cost estimate of a utility double-alkali FGD system is given in a TVA report.[37] The system is a new 500-MW midwestern power unit with a heat rate of 9000 Btu/kWh burning 3.5%-sulfur coal. Sulfur removal was based on meeting the allowable emission of 1.2 lb $SO_2/10^6$ Btu heat input,

Table 9.3 Comparative Costs of Double-Alkali and Other FGD Processes[a]

FGD Process	Investment Requirement ($/kW)	First-Year Revenue Requirement (mill/kWh)
Double-alkali	101	4.19
Lime	90	4.25
Limestone	98	4.02
Magnesia/sulfuric acid	132	5.08
Wellman-Lord/sulfuric acid	131	5.11
Wellman-Lord/sulfur (by Resox)	138	6.44
Wellman-Lord/sulfur (by Allied)	141	6.03
Citrate	143	5.94

[a]Based on a 500-MW midwestern power plant burning 3.5%-sulfur coal; 78.5% SO_2 removal; 30-yr operating life; mid 1979 dollars for capital investment, and mid 1980 dollars for the first-year revenue requirement.

Source: Ref. 5.

equivalent to 78.5% SO_2 removal. The estimated capital investment cost for the system is $50,551,000 -- which corresponds to about $101/kW based on 500-MW gross peak capacity. Note that these estimates include direct and indirect investments, allowance for startup and modifications, interest during construction, land cost, and working capital. Detailed listing of these costs is given in Table 9.4. The annual revenue requirements of the same FGD system total $14,676,000 (4.19 mill/kWh). These include direct costs (material, labor, maintenance) and indirect costs (depreciation, tax, overhead), which are itemized in Table 9.5. These estimates are based on 30-yr plant life, 7000 h/yr operating time for the power unit, onsite solid-waste disposal, and stack gas reheat to 175°F.

9.3.2 Potential Applications

The double-alkali FGD processes appear to be applicable for desulfurizing any flue gas from fossil-fueled boilers, industrial or utility, new or retrofit. These processes have been demonstrated in the U.S. or overseas for both low- and high-sulfur coals. They have the following advantages and disadvantages.

Advantages

1. Capital and operating costs are relatively low. The process utilizes commonly available chemical processing equipment. Materials used are also available in large quantities.
2. Relatively high SO_2 removal efficiencies can be obtained.

Table 9.4 Summary of Estimated Capital Investment for a Sodium-Sulfite-Based Double-Alkali System[a]

Capital Expenditure	Investment ($)	% of Total Direct Investment
Direct Investment		
Materials handling (conveyors, elevators, bins, feeders)	1,710,000	6.4
Feed preparation (feeders, slakers, tanks, agitators, pumps)	833,000	3.1
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,248,000	15.9
SO_2 absorption (four tray towers including presaturator and entrainment separators, recirculation tanks, agitators, pumps)	9,206,000	34.4
Stack gas reheat (four indirect steam reheaters)	1,282,000	4.8
Solids separation (thickener, drum filters, tanks, agitators, pumps, conveyor)	2,352,000	8.8
Solids disposal (onsite disposal facilities, including reslurry tank, agitator, slurry-disposal pumps, and pond-water-return pumps)	1,247,000	4.7
Subtotal	21,235,000	79.4
Services, utilities, miscellaneous	1,274,000	4.8
Total process areas excluding pond construction	22,509,000	84.2
Pond construction	4,241,000	15.8
Total direct investment	26,750,000	100
Indirect Investment		
Engineering design and supervision	1,444,000	5.4
Architect and engineering contractor	331,000	1.2
Construction expense	3,746,000	14
Contractor fees	1,167,000	4.4
Total indirect investment	6,688,000	25
Contingency	6,688,000	25
Total fixed investment	40,126,000	150
Other Capital Charges		
Allowance for startup and modifications	3,589,000	13.4
Interest during construction	4,815,000	18
Total depreciable investment	48,530,000	181.4
Land	837,000	3.1
Working capital	1,184,000	4.4
TOTAL CAPITAL INVESTMENT:	50,551,000	188.9

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; onsite solids disposal.

Midwest plant location represents project beginning mid 1977, ending mid 1980.

Average cost basis for scaling, mid 1979.

Stack gas reheat to 175° F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Disposal pond located 1 mi from power plant.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

Source: Ref. 37.

Table 9.5 Summary of Average Annual Revenue Requirements for a Sodium-Sulfite-Based Double-Alkali System[a]

Annual Revenue Requirement	Annual Quantity	Unit Cost ($)	Total Annual Cost ($)	% of Average Annual Revenue Requirements
Direct Costs				
Raw materials				
Lime (ton)	63,600	42	2,672,200	18.20
Soda ash (ton)	6,060	90	545,400	3.72
Total raw materials cost			3,216,600	21.92
Conversion costs				
Operating labor and supervision (man-hour)	34,500	12.50	431,000	2.94
Utilities				
Steam (10^6 Btu)	489,800	2	979,600	6.67
Process water (kgal)	241,500	0.12	29,000	0.20
Electricity (kWh)	29,100,000	0.029	843,900	5.75
Maintenance				
Labor and material	--	--	1,027,600	7.00
Analyses (man-hour)	4,560	17.00	77,500	0.53
Total conversion costs			3,388,900	23.09
Total direct costs			6,605,500	45.01
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			2,911,800	19.84
Average cost of capital and taxes at 8.6% of total capital investment			4,347,400	29.63
Overheads				
Plant, 50% of conversion costs less utilities			768,200	5.23
Administrative, 10% of operating labor			43,100	0.29
Total indirect costs			8,070,500	54.99
TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS:			14,676,000	100

	mill/kWh	$/ton coal burned	$/$10^6$ Btu heat input	$/short ton S removed
Revenue requirement per unit of input or output	4.19	9.78	0.47	419.31

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; onsite solids disposal.
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit onstream time, 7000 h/yr.
Coal burned, 1,500,100 ton/yr; 9000 Btu/kWh.
Stack gas reheat to 175°F.
S removed, 35,000 short ton/yr; solids disposal 142,750 ton/yr Ca solids including only hydrate water.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $26,750,000; total depreciable investment, $48,530,000; and total capital investment, $50,551,000.
All tons shown are 2000 lb.

Source: Ref. 37.

3. The soluble product in the absorber minimizes erosion problems and the buildup of solids, offering the potential for high system availability. Corrosion and erosion problems could be minor compared to those in wet lime/limestone processes.
4. Since particulates are removed in the prescrubbing stage, the process can remove both SO_2 and particulates.
5. Absorption takes place in the scrubber in an unsaturated solution. The absorbent, therefore, has a large absorption capacity.
6. A relatively low liquid-to-gas ratio is required in the scrubber.
7. High fly ash content can be tolerated in the system.
8. It has a good potential for retrofit application because of smaller process equipment.

Disadvantages

1. Produces large quantities of waste sludge, which requires disposal. Presence of soluble sodium salts in filter cake causes possible environmental concerns.
2. Design complexities must be introduced in order to deal with the following problems:
    a. Necessity of preventing excessive purge of Na_2SO_4 produced as a result of oxidation (Na_2SO_4 is difficult to regenerate).
    b. Necessity of avoiding scrubbing with the clear liquor saturated with calcium sulfate. Excessively high levels of calcium sulfate could lead to scaling problems.
3. Requires some makeup to replenish sodium losses.
4. A wet process, therefore requiring stack reheat.

9.4 PRESENT STAGE OF DEVELOPMENT

9.4.1 Current Applications

The double-alkali process is relatively well developed when compared to other advanced FGD processes. Demonstration-scale double-alkali units are operating in the U.S. with utility boilers, and commercial-scale units are operting with a number of industrial boilers.

Utility (Demonstration Status)

There are three full-scale utility double-alkali systems presently operating in the U.S. The owners of these systems, the size of the systems, and the vendors providing them are: (1) Central Illinois Public Service Co., Newton No. 1, 575-MW unit, Environtech; (2) Louisville Gas and Electric Co., Cane Run No. 6, 277-MW unit, Combustion Equipment Associates/Arthur D. Little, Inc.; and (3) Southern Indiana Gas and Electric Co., A.B. Brown No. 1, 250-MW unit, Food Machinery Corp. All of these systems are of "concentrated" type and they service boilers firing high-sulfur coal.

Brief description of these full-scale double-alkali facilities, their design basis, and reported cost were presented in the EPA 1979 Symposium on Flue Gas Desulfurization.[38] The process expectations and guarantees offered by the vendors for these three systems are summarized in Table 9.6.

The initial operating experience with the double-alkali systems at Southern Indiana Gas and Electric Company's (SIGECO's) A.B. Brown Station Unit #1 boiler, and Louisville Gas and Electric Company's (LG&E's) Cane Run Unit #6 boiler was reported at the EPA 1980 Symposium on Flue Gas Desulfurization. During the first 13 months of routine operation, beginning in August 1979, sulfur dioxide removal of more than 90% was routinely demonstrated at the SIEGECO double-alkali unit. Overall operating cost on an annual-revenue-requirements basis were close to the original projections. The system availability during the period was estimated at 90% and the forced outage rate at 3.3%, which is three times more than that is desired. The FGD system's longest continuous run was 54 days, and it ran 71 days between partial forced outages. The causes of system outages were primarily mechanical and were believed to be correctable in future installations. Engineering system improvements are being made on the SIGECO double-alkali in four unit areas: (1) filter-cake quality, (2) underflow pump recirculation, (3) duplicate lime transfer, and (4) pH controls.[39]

The results from routine operation and the acceptance test conducted on the LG&E's double-alkali unit indicate a greater than 90% SO_2 removal efficiency and system availability of more than 99%. Consumption of raw materials and power were less than guaranteed. Most of the problems initially encountered were mechanical in nature and have been solved or greatly reduced. Further investigation of filter operation, reactor operation, filter cloths, construction materials, and major process component characterization was underway.[40]

Industrial (Commercial Status)

A status summary of full-scale industrial applications of this technology in the U.S. appears in Table 9.7. It shows a total of approximately 640 MW of operating and planned double-alkali capacity in industrial applications. Capacity totaling approximately 265 MW was operational in 1979.

9.4.2 Problems

Problems that can adversely affect the reliability of the technology can be classified as mechanical and chemical.

Table 9.6 Design Parameters for Double-Alkali Systems Being Demonstrated at Three Utilities

Design Parameter	CIPSCO[a,b] Vendor: Environtech	LG&E[c,d] Vendor: CEA/ADL	SIEGECO[e,d] Vendor: FMC
Availability (%)	90% (for 70% load factor over 30-yr life span)	90% (for a 1-yr operating period)	95% (for 30- and 60-d consecutive test runs)
SO_2 removal (% or maximum ppm or lb/10^6 Btu)	90% or outlet SO_2 less than 200 ppm, whichever is greater	200 ppm in scrubber outlet or 95% removal if sulfur content of the coal is greater than 5% by wt	1.2 lb/10^6 Btu (85% for 4.5%-sulfur coal)
Particulate matter removal (lb/10^6 Btu)	≤ 0.10[f]	≤ 0.10 and no net addition of particulate	≤ 0.10 and no net addition of particulate
HCl removal (%)	90%	Not applicable	Not applicable
Sodium consumption (mol Na_2CO_3/mol SO_2)	0.024 (and 0.023 per mol acid gas)	0.045 when maximum coal chloride level is 0.06% (0.5 mol additional for each mol of chloride in the coal above the 0.06% level)	0.03 plus Na lost with chloride
Calcium consumption (mol Ca/mol SO_2 removed)	1.10	Maximum 1.05	Approximately 1
Energy consumption (% of operating rate)	None	1.2% of peak rate (300 MW)	Less than 1%
Solids quantity (% by weight)	None	Minimum 55 (insoluble solids)	Minimum 55

[a] Central Illinois Public Service Co., Newton #1 unit.

[b] Figures for this unit represent expected design values.

[c] Louisville Gas and Electric Co., Cane Run #6 unit.

[d] Figures represent guaranteed design values.

[e] Southern Indiana Gas and Electric Co., A.B. Brown #6 unit.

[f] FGD system designed for possible ESP upsets.

Source: Ref. 38.

Table 9.7 Full-Scale U.S. Industrial Applications of Sodium-Sulfite-Based Double-Alkali Wet-Scrubbing Systems

Process Developer	Precipitant	User	Plant Site	Unit Rating (MW)	Type of Plant[a]	Year of Completion[b]
Food Machinery Corp. (FMC)	Na_2SO_3, $Ca(OH)_2$	FMC	Modesto, Calif.	10 (gas rate)	Reduction kiln^R	1971
General Motors Corp.	$NaOH/Na_2SO_3$, $Ca(OH)_2$	General Motors Corp.	Parma, Ohio	32 (gas rate) 40 (regen.)	Industrial boiler^R	1974
Zurn Industries	$NaOH/NA_2SO_3$, $Ca(OH)_2$	Caterpillar Tractor Co.	Joliet, Ill.	20-30	Industrial boiler^R	1974
FMC	Na_2SO_3, $Ca(OH)_2$	Firestone Tire & Rubber Co.	Pottstown, Penn.	3	Demonstration^R	1975
FMC	Na_2SO_3, $Ca(OH)_2$	Caterpillar Tractor Co.	Mossville, Ill.	50	2 Industrial boilers^{1N,1R}	1975
Zurn Industries	$NaOH/Na_2SO_3$, $Ca(OH)_2$	Caterpillar Tractor Co.	Morton, Ill.	12	Industrial boiler^R	1978
FMC	Na_2SO_3, $Ca(OH)_2$	Caterpillar Tractor Co.	East Peoria, Ill.	100	4 Industrial boilers^{2N,2R}	1978
FMC	Na_2SO_3, $Ca(OH)_2$	Caterpillar Tractor Co.	Mapleton, Ill.	140	Industrial boiler^N	(1979)
FMC	Na_2SO_3, $Ca(OH)_2$	Arco/Polymers, Inc.	Monaca, Penn.	100	Industrial boiler^R	(1980)
FMC	Na_2SO_3, $Ca(OH)_2$	Chanslor Western Oil & Dev. Co.	Bakersfield, Calif.	25	Industrial boiler^N	(1979)
c	d	Dupont, Inc.	Georgia, Ala.	100	Industrial boiler^N	(1987)
c	d	Grissom Air Force Base	Bunker Hill, Ind.	12	3 Industrial boilers^{1N,2R}	(1979)

[a] N = new; R = retrofit.

[b] Dates in parentheses: projected start-up dates.

[c] Vendor not selected.

[d] Data not available.

Source: Ref. 38.

Mechanical Problems

These problems include malfunctioning instruments and malfunctioning mechanical and electrical equipment (e.g., pumps, filters, centrifuge, and valves). Mechanical problems in a commercial FGD system can be minimized by careful selection of construction materials and equipment and by providing spares for equipment items such as pumps and motors that are expected to be in continuous operation. Such operation makes the items prone to failure after a relatively short period of operation. Another important consideration in minimizing mechanical problems is the institution of a good preventive maintenance program.

Chemical Problems

These include scaling, production of poor-settling solid wastes, and excessive buildup of sulfate and limestone used in regeneration. Each factor is associated with reliability of system operation or production of an environmentally acceptable solid waste.

Scaling.

Although one of the primary reasons for developing double-alkali processes was to circumvent the scaling problems associated with lime/limestone wet-scrubbing systems, scaling problems in double-alkali systems have not been totally avoided. Both gypsum and carbonate scale buildup has been recognized in scrubbers. Gypsum scaling occurs when supersaturation is excessive. It can be avoided by maintaining the scrubbing solution unsaturated with respect to gypsum. Carbonate scaling usually occurs as a result of localized high-pH scrubbing liquor being used. This problem can be eliminated by careful control of pH in the scrubber.

Solid Quality.

Under certain conditions, the waste solids produced in the regeneration reaction of double-alkali systems have a tendency not to settle in the scrubber liquor. This creates problems in the operation of thickeners, clarifiers, filters, and centrifuges. The problem is not completely understood, but it is thought to be affected by factors such as reactor configuration, concentrations of soluble sulfate, magnesium and iron in the liquor, concentration of suspended solids in the reaction zone, and use of lime rather than limestone for reaction.

Sulfate Removal.

In double-alkali systems, oxidation reactions result from the conversion of active to inactive sodium. Most of the latter is in the form of sulfate (e.g., Na_2SO_4). Removal of sulfate from the process system is an absolute necessity: failure to do so can cause precipitation and scaling, as well as a deterioration of SO_2 removal capability. Several techniques allow the sodium sulfate to be purged from the system in the liquor, which is abundant in the wet solid-waste product. The resultant solid wastes containing soluble sodium compounds can cause water-pollution problems. Runoff from solid-waste storage can pollute surface water. Leaching and percolation of leachate into the soil can contaminate groundwater.

Limestone For Regeneration. One significant chemical problem yet to be resolved relates to the use of limestone for regeneration. The problem is that the sulfate, magnesium, and iron solids in the scrubbing liquor do not have good settling characteristics in all concentrations. These components tend to reduce the rate of reaction of limestone with the sodium scrubbing solution; their solid properties also tend to deteriorate as the limestone reaction rate decreases.[31]

9.5 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT

The U.S. EPA has been actively involved in the development of double-alkali FGD technology since 1971. EPA's involvement in development of the technology ranges from laboratory tests to full-scale utility demonstrations of the process.

EPA's initial laboratory work was concerned with determining process feasibility and regeneration chemistry. After that, EPA funded a pilot-plant program which was subsequently expanded to include a 20-MW prototype testing at the Scholz Plant of Gulf Power Co. Current activity at this facility involves preparation for testing of limestone as a regenerant. In addition, EPA has funded a program at a General Motors plant to evaluate the full-scale dilute mode of double-alkali scrubbing.

The full-scale demonstration with the Louisville Gas and Electric unit (277 MW) was co-funded by EPA. The program consists of four phases:[38] (1) process design and cost estimation; (2) engineering design, construction, and mechanical testing; (3) startup and acceptance test; and (4) one-year operation and long-term testing. Phases I and II have been completed, and Phase III is in progress. Depending on the level of future funding for EPA, it may expand the current program at LG&E to allow testing of: (1) limestone as an alternative to the more expensive, energy-intensive lime reactant; (2) methods aimed at upgrading the quality of sludge produced and comparing disposal options; and (3) strategies for controlling the multimedia environmental impact of all effluents and emissions.

In addition, EPA may initiate process evaluation programs on the other two full-scale utility applications of the technology (see Table 9.6). Parellel test programs may be conducted to evaluate, characterize and compare these full-size facilities, and allow evaluation of the promising FGD technology.[38]

10 DOWA DOUBLE-ALKALI WET SCRUBBING

SUMMARY

The DOWA double-alkali FGD process (also called the DOWA aluminum-sulfate-limestone process) uses a basic aluminum-sulfate solution as the absorbent and limestone as the precipitant to remove SO_2 from flue gases. This process was developed by DOWA Mining Co. in Japan. It has been commercially applied in Japan to desulfurize flue gas from an oil-fired boiler and waste gases from smelters, roasters, and sulfuric-acid plants. Its applicability to coal-fired boilers is now being tested in the U.S. The SO_2 removal efficiency of the DOWA process appears to be greater than 90%. The SO_2 removed in the process is converted into gypsum which can either be sold or disposed of on land. The process is simple and easy to operate. Capital and operating costs, based on a preliminary estimate, are competitive with those of the lime and limestone processes. More study is needed to evaluate process design parameters, optimize process performance, and determine the environmental effects. Furthermore, economic evaluation of the process needs to be updated.

10.1 TECHNICAL ASPECTS

10.1.1 Process Description[41]

The DOWA process is a dual-alkali wet-scrubbing process that utilizes a basic aluminum-sulfate solution to absorb the SO_2 and a limestone solution to regenerate the absorbent. The process was developed by the Dowa Mining Co. of Tokyo, and it will be marketed in the United States by the Air Correction Division of UOP, Inc. The process is now in commercial operation in Japan at an oil-fired boiler, smelters, and sulfuric-acid plants. The Shawnee prototype DOWA installation is the first test of the DOWA process with flue gas from a coal-fired boiler.

Figure 10.1 is a flow diagram of the Shawnee prototype DOWA installation. The entire process can be subdivided into four sections: absorber, oxidizer, neutralizer, and solid dewaterer. Sulfur dioxide absorption occurs in the absorber. The oxidation process step occurs in both the absorber and the absorber hold tank. A bleedstream of absorbent is pumped to the neutralizer tanks, where the limestone required for neutralization is added. The neutralizer product overflows from the second neutralizer into a conventional thickener. The thickener overflow is collected in the reclaimed absorbent tank, and the thickener underflow is pumped to the filter for final dewatering of the gypsum by-product. The filtrate is returned to the reclaimed absorbent tank. A portion of the thickener underflow is recycled to the first neutralizer tank to provide gypsum seed crystals for the neutralization and gypsum-precipitation step.

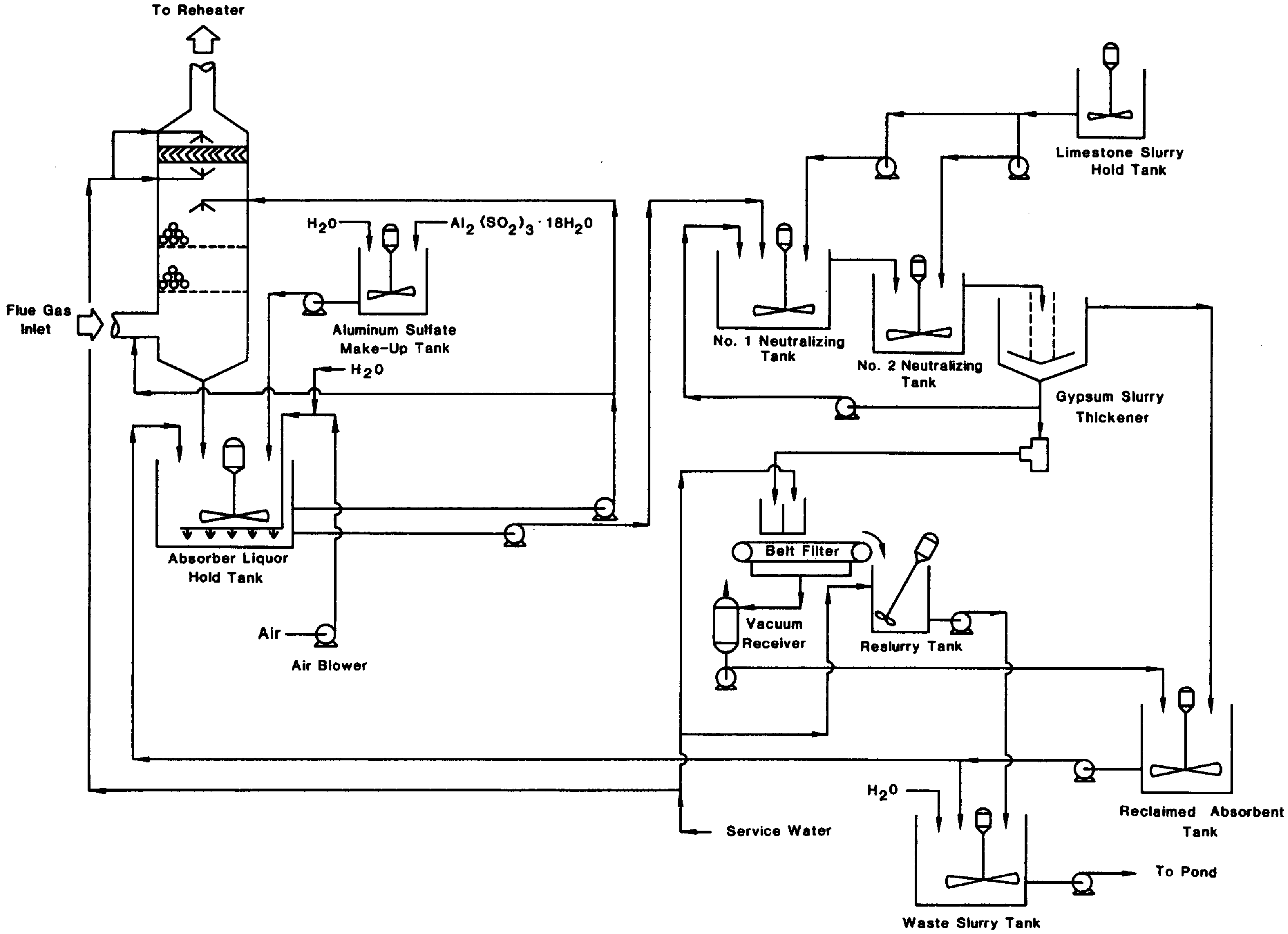


Fig. 10.1 Flow Diagram of DOWA Double-Alkali Wet-Scrubbing FGD Process (Source: Ref. 41)

10.1.2 Process Chemistry[41]

The overall chemical reactions in each of the major process steps are:

- Absorption: $$Al_2(SO_4)_3 \cdot Al_2O_3 + 3SO_2 \rightarrow Al_2(SO_4)_3 \cdot Al_2(SO_3)_3 \tag{10.1}$$
- Oxidation: $$Al_2(SO_4)_3 \cdot Al_2(SO_3)_3 + 3/2O_2 \rightarrow Al_2(SO_4)_3 \cdot Al_2(SO_4)_3 \tag{10.2}$$
- Neutralization: $$\begin{aligned} & Al_2(SO_4)_3 \cdot Al_2(SO_4)_3 + 3CaCO_3 + 6H_2O \\ & \rightarrow Al_2(SO_4)_3 \cdot Al_2O_3 + 3CaSO_4 \cdot 2H_2O + 3CO_2 \end{aligned} \tag{10.3}$$

The process is more accurately defined by the principal intermediate steps within the process steps, as follows.

- Absorption

$$SO_2(g) \rightarrow SO_2 \text{ (diss.)} \tag{10.4}$$

$$SO_2 \text{ (diss.)} + H_2O \rightleftarrows H_2SO_3 \text{ (diss.)} \rightleftarrows H^+ + HSO_3^- \tag{10.5}$$

In addition to the preceding reactions, the following reactions, which describe the buffering action of basic aluminum sulfate, are important.

$$Al^{+3} + xOH^- \rightleftarrows Al(OH)_x^{+(3-x)} \tag{10.6}$$

$$Al(OH)_x^{+(3-x)} + H^+ \rightleftarrows H_2O + Al(OH)^{+(4-x)}_{(x-1)} \tag{10.7}$$

$$Al_2(SO_4)_3 \rightarrow 2Al^{+3} + 3SO_4^{-2} \tag{10.8}$$

- Oxidation

$$H_3O_3^- + 1/2O_2 \rightarrow H^+ + SO_4^{-2} \tag{10.9}$$

- Neutralization

$$SO_4^{-2} + CaCO_3(s) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \downarrow + CO_3^{-2} \tag{10.10}$$

$$CO_3^{-2} + H_2O \rightleftarrows HCO_3^- + OH^- \tag{10.11}$$

$$HCO_3^- + H_2O \rightleftarrows H_2CO_3 + OH^- \tag{10.12}$$

$$H_2CO_3 \text{ (diss.)} \rightleftarrows H_2O + CO_2 \text{ (diss.)} \rightleftarrows CO_2(g) \tag{10.13}$$

The last reaction goes to completion at pH3.

In summary, sulfur dioxide is absorbed in a solution of basic aluminum sulfate at a pH of approximately 3 [reactions (10.4) through (10.8)]. The resultant sulfite in the liquor is oxidized to sulfate by oxygen in the flue gas and in the air which is sparged into the liquor [reaction (10.9)]. The oxidized liquor is regenerated to basic aluminum sulfate by neutralization with limestone [reactions (10.10) through (10.13)]. The gypsum by-product from the neutralization step is removed by gravitational settling and filtration. The filtrate and clarified liquor are returned to the process.

Highly efficient SO_2 removal by the process requires that the equilibrium of reaction (10.5) be shifted to the right to allow more HSO_3^- in solution. This is accomplished by more efficient oxidation of the absorber liquor [reaction (10.9)].

The concentrations of chloride and magnesium in the process liquor are controlled by a purge stream. The aluminum content of the purged liquor is recovered by adding excess limestone to precipitate the aluminum as aluminum hydroxide. The precipitated aluminum is separated from the supernatant and returned to the process.

10.1.3 Process Performance

SO_2 Removal Efficiency. Table 10.1 shows the performance data from three existing commercial DOWA systems in Japan, indicating that this FGD system can remove more than 90%, and in several cases, greater than 95% of SO_2 from waste gases.[42] Performance of the process has been improved by adding a small amount of soluble metallic catalyst to the absorbent. The test results from Naikai Engyo (Table 10.1) indicate that catalytic action has led to lowering the liquid-to-gas ratio while maintaining the SO_2 removal efficiency. Catalytic action also helped in promoting oxidation reaction and reducing the amount of air to the process.

DOWA double-alkali process tests at the Shawnee test facility were the first application of the DOWA process to flue gas from a coal-fired boiler. The initial tests utilizing the existing turbulent contact absorber (TCA)

Table 10.1 Performance of Four Existing DOWA Double-Alkali FGD Systems

Performance Characteristic	Taenaka Mining (Kiln)	DOWA Mining (H_2SO_4 Plant)	Naikai Engyo, #1 (Industrial Oil-Fired Boiler)	Naikai Engyo, #2 (Industrial Oil-Fired Boiler)
Liquid-to-gas ratio				
(L/normal m^3)	5	2.5	8	5
(gal/1000 scf)	35	17.5	56	35
Inlet SO_2 (ppm)	7500	400-700	1500-1700	1500-1700
SO_2 removal efficiency (%)	>95	>92.5	>95	>95
Remark		w/catalytic actions		w/catalytic actions

Source: Ref. 42.

resulted in a maximum SO_2 removal efficiency of 85-90%. After switching to a fixed-bed packed absorber, which provided an improved gas-flow distribution and gas-liquor content, the SO_2 removal efficiency of the process was improved, to more than 90%.[41]

System Reliability. Existing commercial DOWA FGD units in Japan have been in smooth, continuous operation since their startup. There seem to be no appreciable process-related problems. The DOWA FGD unit at Okayama Plant registered a 91% availability (hours of FGD operation over total number of hours in the period), and a 99.6% operability (hours of FGD operation over hours of boiler operation) during the period from August 1976 to July 1977.[43] Soft deposits of solids on the packing in the scrubber unit caused appreciable increase of gas pressure drop, resulting in operation difficulties. The problem was solved when the deposits were removed during the annual maintenance shutdown.

Extensive reliability tests were not conducted at the Shawnee facility DOWA process tests. However, the report indicates that no significant reliability problems were identified. There was no scale formation in the absorber, and high concentration of fly ash in the process absorbent did not affect the process performance.[41]

10.1.4 Water, Chemical, and Power Requirements

Operation of a DOWA system requires material inputs including aluminum sulfate, catalyst, air calcium carbonate (limestone), and water. Table 10.2 shows the operating requirements for the DOWA system at the Okayama plant.

Electricity required to run the DOWA system could derate the generation capacity by nearly 2%. Rates of material requirements are estimated to be: 0.058 kg aluminum sulfate, 0.0065 kg catalyst, 2.86 kg limestone, and 35.8 kg water -- for every kg of sulfur removed.[42]

Table 10.2 Operating Requirements of the DOWA FGD System at the Okayama Plant, Naikai Engyo, Japan[a]

Requirement	Amount
Electricity (kW)	480
Process water (ton/h)	5.5
Aluminum sulfate (kg/h)	9
Catalyst (kg/h)	1
$CaCO_3$ (ton/h)	0.44
Labor (men/shift)	1

[a] Flue gas volume, 7200 normal m^3/h; sulfur in fuel oil, 3%; SO_2, 1500 ppm; SO_2 removal, 92%; rate of operation, 95%, 347 d/yr; by-product, $CaSO_4 \cdot 2H_2O$ 0.8 ton/h; generating capacity, 25 MW.

Source: Ref. 42.

10.2 ENVIRONMENTAL IMPLICATIONS

10.2.1 Gaseous Emissions

As mentioned earlier, the DOWA process can achieve more than 90% SO_2 removal efficiency. No NO_x removal has been shown for the process.

10.2.2 Liquid Effluents

Under normal conditions, no wastewater is generated in the process. However, recirculation of absorbent in the system could build up concentration of impurities. One of the impurities is chlorine, which is derived from fuel and process water. The accumulation of chlorine causes corrosion. Magnesium is also an impurity. It is derived from limestone and interferes with the reaction of sodium sulfite and limestone.

To reduce these impurities, some wastewater will have to be purged. At the Okayama plant application, no wastewater was purged during the first year of operation, resulting in a buildup of chlorides to 2000 ppm and magnesium to 20,000 ppm in the scrubber liquor. Since then about 8 ton/d of wastewater have been purged to prevent further accumulation of impurities. Before being discharged, the purged wastewater was treated by neutralization with lime to precipitate aluminum hydroxides and the metallic catalyst. The precipitates are returned to the scrubber system.

10.2.3 Solid By-Product

The technology produces a solid by-product, gypsum. The gypsum produced from commercial DOWA facilities contains a small amount (0.05%) of aluminum and could also contain fly ash if coal flue gas is treated.

The composition of a typical gypsum by-product from tests at the Shawnee facility is given in Table 10.3.[41] Nearly complete oxidation of sulfite to sulfate was observed. A relatively high aluminum (Al) concentration in the by-product gypsum was attributed to the inadequate cake-washing procedures. Lower Al concentration could be expected with improved cake washing, such as the 0.05% concentration achieved in commmercial facilities in Japan.

In Japan, gypsum produced from the DOWA system has been sold and used for wallboard production and as a retarder for cement setting. The small amount of aluminum does not affect the quality of wallboard or cement produced from the gypsum. Because of potential oversupply of gypsum, an investigation

Table 10.3 Composition of By-Product Gypsum, the DOWA Shawnee Test

Component	% by Wt (Dry)
Aluminum	0.3
Calcium	21.8
Carbonate	nil
Sulfite	--
Sulfate	53.8
Total solids (wet basis)	81.8
Acid insolubles	nil

Source: Ref. 41.

of possible new uses of gypsum has been underway. Its application as a building material, either by itself or reinforced with polymer to form a gypsum/plastic composite, appears to be promising. If a market is not available, gypsum can be disposed of on land. Environmental effects of gypsum from the DOWA process are not entirely known, however.

10.3 ECONOMIC AND MARKET FACTORS

10.3.1 Operating and Capital Costs

Capital investment and annual revenue requirements of the DOWA double-alkali wet scrubbing were recently estimated by the Tennessee Valley Authority (TVA).[44] Considering the status of the technology, these estimates are preliminary and should be treated as such. The estimates are based on a 500-MW new coal-fired power unit, burning 3.5%-sulfur coal, emitting 1.2 lb $SO_2/10^6$ Btu heat input (or, 78.5% SO_2 removal), and generating gypsum solids to be disposed of on site. The costs in Tables 10.4 and 10.5 reflect these assumptions. Total capital investment (1979 dollars) was estimated at $49,454,000 ($99/kW). Total annual revenue requirements were estimated at $15,253,100 (4.36 mill/kWh).

Based on the same TVA study, capital and operating requirements of the DOWA system are quite comparable with those of the lime and limestone systems, but lower than those of such other advanced FGD systems as MgO and Wellman-Lord wet-scrubbing process. A comparison of costs between DOWA system and other FGD systems is presented in Table 10.6.

10.3.2 Potential Applications

Based on the commercial applications in Japan and the prototype tests in the U.S., it is recognized that the DOWA double-alkali process may be applied to any flue gas containing SO_2. The process is simple, and the plant is easy to operate. Chemicals including limestone and aluminum sulfate are readily available and do not adversely affect environment. An SO_2 removal efficiency of more than 90% can be achieved. DOWA wet scrubbing also has the following advantages and disadvantages:

Advantages

- Scrubbing solution is used, thus minimizing scaling and plugging problems.
- Regeneration of absorbent is done with limestone, which is considerably cheaper than lime.
- End product is in the form of gypsum, which has market value, is easy to handle, and creates little disposal problem.
- Capital and operating requirements are comparable to those of limestone and lime FGD systems and lower than those of some other advanced FGD systems.

Table 10.4 Summary of Estimated Capital Investment for a DOWA Double-Alkali Wet-Scrubbing System[a]

Capital Expenditure	Investment ($)	% of Total Direct Investment
Direct Investment		
Materials handling (hoppers, feeders, conveyors, elevator, bins, shaker, puller)	2,095,000	8.1
Feed preparation (feeders, crushers, ball mills, hoist, tanks, agitators, pumps)	1,557,000	6
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to reactor, exhaust gas ducts, dampers from reactor to reheater and stack)	4,318,000	16.7
SO_2 absorption (four TCA scrubbers including presaturators and entrainment separators, recirculation tanks, agitators, pumps)	8,937,000	34.5
Stack gas reheat (four indirect steam reheaters)	1,282,000	4.9
Oxidation (towers and compressors)	3,070,000	11.9
Magnesium purge (tanks, agitator, and pumps)	78,000	0.3
Neutralization (tanks, agitators, and pumps)	667,000	2.6
Solids disposal (thickener, filters, pumps, conveyors, and mobile equipment)	2,232,000	8.6
Subtotal	24,236,000	93.6
Services, utilities, and miscellaneous	1,454,000	5.6
Total process areas excluding disposal field preparation	25,690,000	99.2
Disposal field preparation	209,000	0.8
Total direct investment	25,899,000	100
Indirect Investment		
Engineering design and supervision	1,604,000	6.2
Architect and engineering contractor	397,000	1.5
Construction expense	3,734,000	14.4
Contractor fees	1,139,000	4.4
Total indirect investment	6,874,000	26.5
Contingency	6,555,000	25.3
Total fixed investment	39,328,000	151.8
Allowance for startup and modifications	3,912,000	15.1
Interest during construction	4,719,000	18.2
Total depreciable investment	47,959,000	185.1
Land	315,000	1.2
Working capital	1,180,000	4.6
TOTAL CAPITAL INVESTMENT:	49,454,000	190.9

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; onsite solids disposal.

Evaluation represents project beginning mid 1977, ending mid 1980. Average cost basis for scaling, mid 1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Disposal pond located 1 mi from power plant.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

Source: Ref. 44.

Table 10.5 Summary of Average Annual Revenue Requirements for a DOWA Double-Alkali System[a]

Annual Revenue Requirement	Annual Quantity	Unit Cost ($)	Total Annual Cost ($)	% of Average Annual Revenue Requirements
Direct Costs				
Raw materials				
Limestone (ton)	110,000	7	770,000	5.05
Aluminum sulfate, 57% (ton)	3,900	185	721,500	4.73
Catalyst (lb)	6,400	0.70	500	-
Total raw materials cost			1,492,000	9.78
Conversion costs				
Operating labor and supervision				
FGD (man-hour)	27,500	12.50	343,800	2.25
Solids disposal (man-hour)	10,400	17	176,800	1.16
Utilities				
Steam (10^6 Btu)	489,800	2	979,600	6.42
Process water (kgal)	193,300	0.12	23,200	0.15
Electricity (kWh)	80,987,900	0.029	2,348,600	15.40
Maintenance				
Labor and material	--		1,547,700	10.15
Analyses (man-hour)	3,220	17	54,700	0.36
Gypsum disposal (ton)	257,800	0.11	28,400	0.19
Total conversion costs			5,502,800	36.08
Total direct costs			6,994,800	45.86
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			2,877,500	18.87
Average cost of capital and taxes at 8.6% of total capital investment			4,253,000	27.88
Overheads				
Plant, 50% of conversion costs less utilities			1,075,700	7.05
Administrative, 10% of operating labor			52,100	0.34
Total indirect costs			8,258,300	54.14
TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS:			15,253,100	100

	mill/kWh	$/ton coal burned	$/$10^6$ Btu heat input	$/short ton S removed
Revenue requirement per unit of input or output	4.36	10.17	0.48	436

[a]Basis:500-MW new coal-fired power unit, 3.5%-sulfur in coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; onsite solids disposal)

Midwest plant location, 1980 revenue requirements.

Remaining life of power plant, 30 yr.

Power unit onstream time, 7000 h/yr.

Coal burned, 1,500,100 ton/yr, 9000 Btu/kWh.

Stack gas reheat to 175°F.

Sulfur removed, 35,000 short ton/yr; solids disposal 142,750 ton/yr Ca solids including only hydrate water.

Investment and revenue requirement for removal and disposal of fly ash excluded.

Total direct investment, $25,899,000; total depreciable investment, $47,959,000; and total capital investment, $49,454,000.

All tons shown are 2000 lb.

Source: Ref. 44.

Table 10.6 Comparative Costs of DOWA Double-Alkali and Other Wet-Scrubbing FGD Processes[a]

FGD Process	Unit Investment Requirement ($/kW)	Revenue Requirement Per Unit of Output (mill/kWh)
DOWA double-alkali	99	4.36
Lime	91	4.25
Limestone	98	4.11
Magnesia (to H_2SO_4)	141	5.24
Wellman-Lord (to H_2SO_4)	143	5.45
Wellman-Lord (to sulfur)	148	6.14

[a] Based on a 500-MW midwestern power plant burning 3.5%-sulfur coal; 78.5% SO_2 removal; 30-yr operating life; 1979 dollars for investment requirements and 1980 dollars for annual revenue requirements. By-product sales revenues are included for magnesia and Wellman-Lord systems.

Source: Ref. 44.

- Conversion from existing lime/limestone systems to DOWA system is relatively easy.
- Oxidation tolerance of absorbent is unlimited.

Disadvantages

- Relatively high liquid-to-gas ratio.
- Limited operating experiences on coal-fired boilers.
- Scrubbed flue gas requires reheat.
- Uses an absorbent of low pH, thus requiring more acid-resistant materials in process equipment.

10.4 PRESENT STAGE OF DEVELOPMENT

10.4.1 Commercial Applications in Japan

The DOWA FGD process was initially developed by DOWA Mining Co., one of the largest manufacturers of nonferrous metals in Japan. Mitsui Shipbuilding Co. has joined DOWA for further development of the process.

At the beginning of 1978, there were seven DOWA double-alkali commercial plants operational in Japan, having a combined total capacity of 200 MW. Table 10.7 summarizes these commercial systems by user, plant capacity, source of waste gas, and year of completion. Only one unit was applied to an industrial boiler; the remaining units were for sintering plants, smelters, and a

Table 10.7 Full-Scale Commercial DOWA Systems in Japan

User	Capacity 1000 normal m^3/h	Capacity MW	Source of Gas	Inlet SO_2 (ppm)	Year of Completion
Taenaka		1	Kiln	7500	1972
DOWA Mining	150 · 2		H_2SO_4 plant	650	1974
Naikai Engyo	70	25	Industrial boiler	1500	1976
Yahagi Iron	50		Sintering plant	2500	1976
Niho Seiko	30		Sintering plant	5000	1976
Kowa Seiko	72		H_2SO_4 plant	750	1978
Mitsubishi Metal	140		Smelting furnace	4000	1978

Source: Ref. 42.

sulfuric-acid plant. These facilities have been operated successfully, according to a recent report.[38]

10.4.2 Prototype Tests on Coal Boilers

In the U.S., the DOWA process has been licensed by the Air Correction Division of UOP, Inc. A series of tests of the DOWA process with flue gas from coal-fired boilers has recently been conducted at the TVA's Shawnee Station, Paducah, Ky. The testing program was jointly funded by TVA, EPRI, and UOP, Inc. The EPA also provided a part of the funding. Continuous operation of the test unit started in Jan. 1980. The unit received a flue-gas stream ranging from 13,000 to 27,000 acfm at 300°F. The program's general objective is to evaluate the applicability of DOWA wet scrubbing to coal-fired boilers. Specific tests were conducted to examine whether fly ash would interfere with the process chemistry. Results from these tests have been highlighted in Secs. 10.1 and 10.2 of this report. More detailed presentation of the test results is available in Ref. 41.

10.5 STATUS OF RESEARCH FOR FURTHER DEVELOPMENT[41]

UOP, Inc., is now performing laboratory studies and installing an integrated pilot plant to further optimize the DOWA process. TVA and UOP have independently proposed further DOWA process tests at Shawnee. These further tests are contingent upon completion of the laboratory and pilot tests and an economic evaluation of the process.

11 CHIYODA-121 WET SCRUBBING

SUMMARY

The Chiyoda FGD process identified as Thoroughbred 121 (or CT-121) is an improved version of conventional lime and limestone wet-scrubbing processes. It uses a compact reactor that combines in one vessel sulfur-dioxide absorption, sulfite oxidation, acid neutralization, and gypsum crystallization. Combining all these processes could greatly simplify the design and operation of an FGD system. Development of this FGD process is relatively recent. Results from limited pilot and prototype tests indicate a high degree of reliability, efficient desulfurization, and ease of operation. Preliminary cost estimates also reveal that the CT-121 process could be economically competitive with lime and limestone scrubbing. A full-scale demonstration by a utility company appears to be a logical next step before commercialization.

11.1 TECHNICAL ASPECTS

11.1.1 Process Description

The Chiyoda Thoroughbred 121 (CT-121) process is an advanced flue-gas-desulfurization technology combining the desirable feature of Chiyoda's CT-101 process and lime/limestone scrubbing processes. It was developed by Chiyoda Chemical Engineering and Construction, Ltd., Yokohama, Japan, and Chiyoda International Corp., Seattle, Wash. The process is still in development stages. Tests have been conducted at laboratory and pilot-scale units in Japan since 1975. Since mid 1978, a 23-MW prototype demonstration plant has been in operation at Gulf Power's Scholz coal-fired steam plant (Sneads, Fla.), and the plant includes a CT-121 system.

A schematic flow diagram of the CT-121 FGD process is presented in Fig. 11.1. The flue gas is at first quenched with water and then introduced directly to a jet-bubbling reactor. Depending on the efficiency of the electrostatic precipitator and marketability of the gypsum by-product, a prescrubber may or may not be needed for fly ash removal. The jet-bubbling reactor is a single vessel consisting of a flue-gas inlet and outlet, air inlet, limestone-slurry inlet, and gypsum-slurry outlet. Mechanical agitation is also provided. The SO_2 in the gas is absorbed, oxidized, and neutralized in this single reactor, where the flue gas is sparged into the absorbent through the distributor to generate sufficient froth for liquid/gas contact. Sulfur dioxide is removed from the gas, producing sulfite, and the sulfite is oxidized to sulfate in the absorbent. Oxidizing air from the bottom of the reactor supplies sufficient oxygen to completely oxidize the sulfite. The clean gas is discharged from the reactor after separation of entrained droplets.

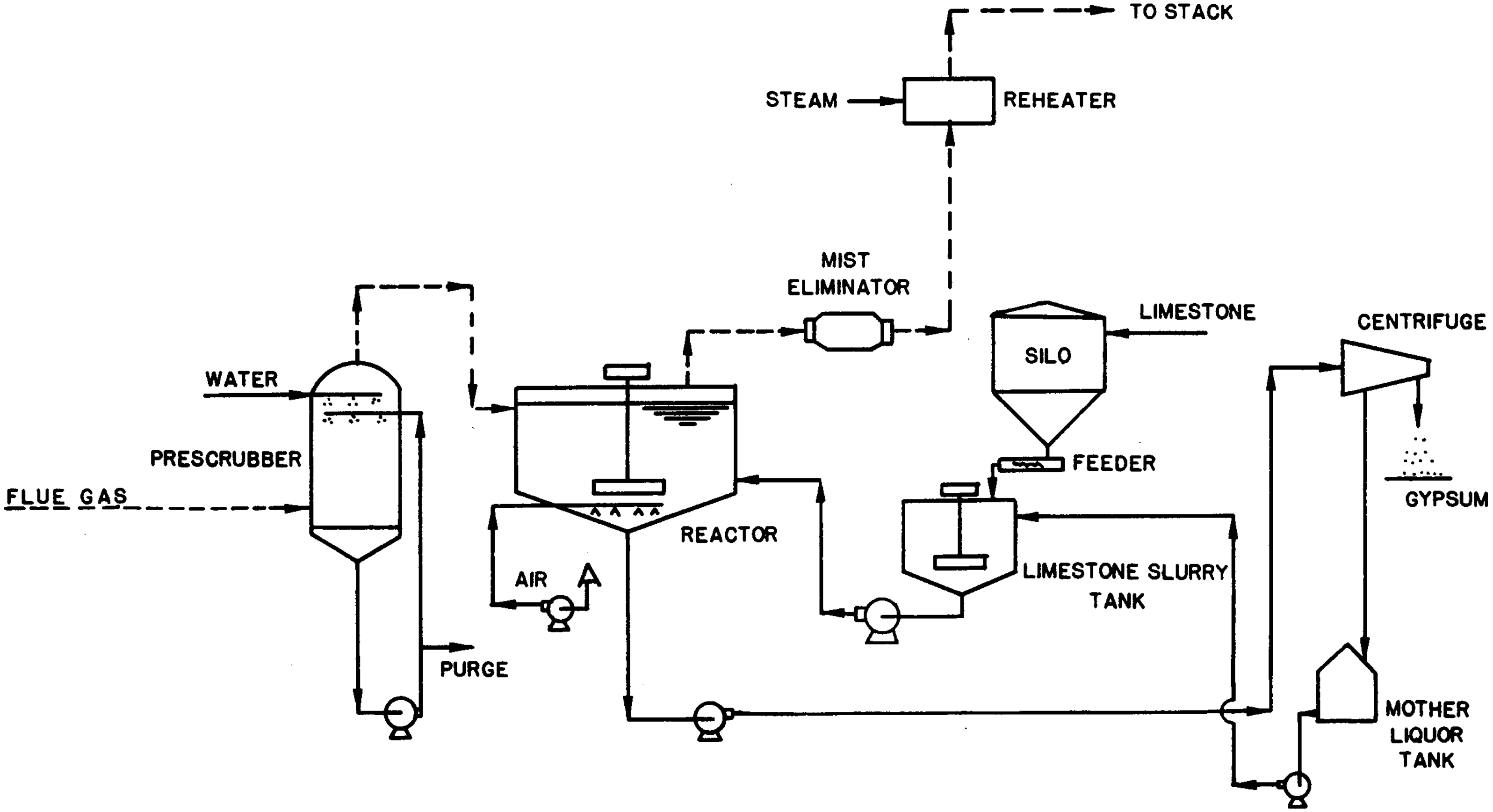


Fig. 11.1 Flow Diagram of Chiyoda-121 Wet-Scrubbing FGD Process (Source: Ref. 48)

Limestone is introduced in the reactor to precipitate the sulfate as gypsum. A homogenous mixture of gas, liquid, and solids in the reactor is maintained by gas and mechanical agitation. The crystallized gypsum by-product is discharged from the reactor as a slurry containing 5% to 20% solids by weight. The solids are separated and the mother liquid is returned to the reactor. Solids separation can be achieved by a variety of means: filtration, centrifugation, or ponding/settling with recycle overflow.

Jet Bubbling Reactor. The jet-bubbling reactor (JBR) is a single vessel composed of two zones: a jet-bubbling zone and a reaction zone. The configuration is shown in Fig. 11.2.

In the jet-bubbling zone, flue gas is sparged into a relatively shallow liquid layer through an array of vertical spargers having their open ends submerged below the liquid surface. High-velocity gas entrains surrounding liquid, creating a jet-bubbling (froth) layer with a gas-liquid interfacial area that provides effective SO_2 removal. The gas-contact time in the jet-bubbling zone ranges from 0.5 to 1.5 seconds.

The reaction zone located at the lower part of the JBR provides moderate agitation by means of air bubbling and mechanical stirring. Oxidizing air is introduced at several times the stoichiometric requirement. The liquid-residence time in the reaction zone ranges from one to four hours.

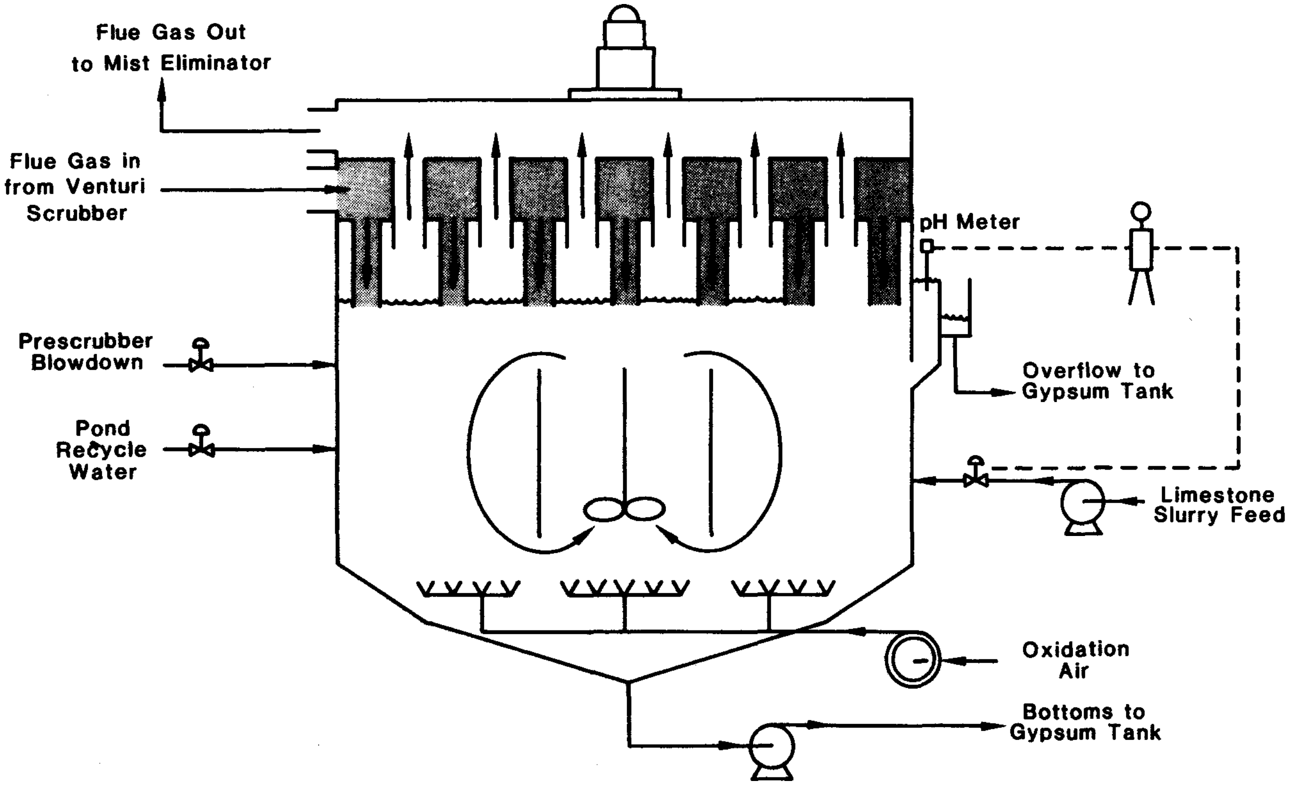


Fig. 11.2 Flow Diagram of the Jet-Bubbling-Reactor Stage of Chiyoda-121 FGD Scrubbing (Source: Ref. 45)

11.1.2 Process Chemistry

The chemistry of the CT-121 process is similar to that of conventional limestone wet scrubbing; it is also different, however, in that SO_2 is oxidized to sulfate, leaving only trace amount of sulfite. It is different from indirect limestone scrubbing processes (e.g., the double-alkali processes) in that all chemical steps are carried out in one vessel.

The overall reaction in the JBR is

$$SO_2 + CaCO_3 + 1/2O_2 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O + CO_2 \tag{11.1}$$

The following reactions occur in the jet-bubbling zone.

$$SO_2\ (g) \leftrightarrows SO_2\ (aq) \tag{11.2}$$

$$SO_2\ (aq) + H_2O \rightarrow H_2SO_3 \tag{11.3}$$

$$H_2SO_3 \leftrightarrows HSO_3 + H^+ \tag{11.4}$$

$$HSO_3^- \leftrightarrows SO_3^= + H^+ \tag{11.5}$$

$$SO_3^= + 1/2O_2(aq) \rightarrow SO_4^= \tag{11.6}$$

$$CaCO_3(s) \rightleftarrows CaCO_3\ (aq) \tag{11.7}$$

$$CaCO_3(aq) + H^+ \rightleftarrows Ca^{++} + HCO_3^- \tag{11.8}$$

$$HCO_3^- + H^+ \rightarrow H_2O + CO_2 \tag{11.9}$$

$$Ca^{++} + SO_4^= + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \tag{11.10}$$

Reactions occuring in the reaction zone are as follows.

$$O_2(g) \leftrightarrows O_2\ (aq) \tag{11.11}$$

$$SO_3^= + 1/2O_2\ (aq) \rightarrow SO_4^= \tag{11.12}$$

$$CaCO_3(s) \leftrightarrows CaCO_3\ (aq) \tag{11.13}$$

$$CaCO_3(aq) + H^+ \rightleftarrows Ca^{++} + HCO_3^- \tag{11.14}$$

$$Ca^{++} + SO_4^= + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \tag{11.15}$$

In the jet-bubbling zone, the controlling steps are gas-phase mass transfer of SO_2, dissolution of $CaCO_3$, and hydration of SO_2 to give H^+. In the reaction zone, liquid-phase mass transfer of O_2 and gypsum crystal growth are the controlling steps.

11.1.3 Process Performance

Particulate and SO_2 Removal Efficiency. The jet-bubbling reactor alone has a particulate removal efficiency of 90%. The CT-121 FGD system combined with other particulate systems can achieve greater fly ash removal efficiency.

Results from pilot and prototype demonstration units (see Table 11.1) indicate that the CT-121 FGD process can achieve an SO_2 removal efficiency of about 90%. Higher removal efficiencies can probably be achieved, but would be at the cost of greater pressure drops and capital cost. Prototype tests at the Scholz coal-fired utility boiler show that three parameters -- the JBR overflow pH, the JBR pressure drop, and the SO_2 concentration at the flue-gas inlet -- influence SO_2 removal efficiency to the greatest extent.[45] These three parameters were fit to a theoretical expression for SO_2 removal efficiency utilizing prototype test results. It resulted in Eq. (11.16), which predicts the removal for inlet SO_2 levels less than 22,000 ppm; and Eq. (11.17), which predicts the efficiency for inlet flue gas at higher SO_2 concentrations.

$$\text{Fractional } SO_2 \text{ removal} = \frac{1 - \exp\left[-3.49\left(\frac{\Delta P}{15.4}\right)^{1.07}\right]}{1 + 56.9\left(\frac{15.4}{\Delta P}\right)(10^{-pH})\left(\frac{SO_2}{1000}\right)^{0.11}} \qquad (11.16)$$

$$\text{Fractional } SO_2 \text{ removal} = \frac{1 - \exp\left[-3.85\left(\frac{\Delta P}{15.4}\right)^{1.44}\right]}{1 + 0.84\left(\frac{15.4}{\Delta P}\right)(10^{-pH})\left(\frac{SO_2}{1000}\right)^{5.26}} \qquad (11.17)$$

where

ΔP = JBR pressure drop (in. of water)

SO_2 = inlet flue-gas sulfur-oxide concentration (ppm)

pH = the pH value measured at the JBR overflow.

These equations show that increases in either pH or ΔP would result in improved SO_2 removal efficiency for two reasons: increased pH results in decreased SO_2 back pressure in the froth zone, and increased ΔP reflects longer gas-liquid contact time and/or more efficient flue-gas sparging. These equations also indicate that the SO_2 concentration has minimal effect on SO_2 removal when the inlet SO_2 concentration remains below about 2200 ppm. Above this level, increases in SO_2 concentration causes a fairly rapid decline in SO_2 removal efficiency -- see SO_2 exponent term in Eq. (11.17).

Several other process parameters were also tested, including JBR configuration, limestone particle size, and chloride concentrations in the JBR slurry -- but their effects on SO_2 removal were relatively minor. The flue-gas flow rate and the oxidation air rate were judged to have a measurable impact on SO_2 removal, but there was not sufficient time to test their impacts.[45]

Table 11.1 Performance of Chiyoda-121 FGD Systems

Process Characteristic	Japanese Pilot Units[a] Low-Sulfur Case	Japanese Pilot Units[a] High-Sulfur Case	U.S. 23-MW Prototype Unit[b]
Fuel	Oil	Oil	Coal
Flue-gas flow rate (scfm)	585	585	--
Inlet SO_2 (ppm)	1100	2900-3000	2000
SO_2 removal (%)	92	87	85-90
Limestone utilization (%)	99.3	99.1	98

[a] Ref. 46.

[b] Refs. 45 and 47.

Scaling Tendency. Sulfate scaling problems are generally found in lime and limestone FGD systems and are a source of potential trouble in CT-121 systems. However, these problems may be prevented in the CT-121 by maintaining a relatively high gypsum-crystal concentration and having sufficient liquid volume. This way, the area for gypsum-crystal growth is increased and the degree of supersaturation decreased. As a result, gypsum precipates only on the surfaces of gypsum crystals, eliminating calcium sulfates deposition, or scaling, on the reactor walls and internals.

Calcium-carbonate and calcium-sulfite scaling may not be a problem for the CT-121 process due to complete oxidation of sulfite, calcium utilization of nearly 10%, and optimally selected pH and reactor capacity.

The solids produced at the Scholz test unit were reported to contain more than 97% gypsum. No sulfite solids were measured. Also, there were only small amounts of calcium carbonate measured in the JBR underflow slurry. According to the Scholz test report, little scale deposition occurred in the JBR. There were some random patches of gypsum scale on various surfaces, but none of the depositions presented threats to system performance.[45]

Process Dependability. The prototype at Scholz tested over a 10-month period was shown to operate reliably under a variety of test conditions. The four performance parameters employed by EPA to measure an FGD system's dependability are presented in Table 11.2. Phase 0 and Phase III test periods involved some planned outages, which penalized both the operability and utilization factors. During Phase I and Phase II, all four factors were high, making the system highly dependable. There were only 22 hours of forced outage during these periods. Of these, 21 hours were due to limestone-feeder problems.[45]

Table 11.2 Chiyoda-121 Performance Characteristics, Measured at Scholz Facility

Test Period	Performance Characteristic (%)			
	Availability[a]	Reliability[b]	Operability[c]	Utilization[d]
Phase 0	99.2	99.1	88	88
Phases I and II	99.3	99.3	97.3	97.3
Phase III	99.5	99.1	58.6	58.6
Total Program Average	99.3	99.2	90	90

[a] Availability: Hours the FGD system is available for operation (whether operated or not) divided by hours in the period.

[b] Reliability: Hours the FGD system was operated divided by the hours the FGD system was called upon to operate.

[c] Operability: Hours the FGD was operated divided by the boiler operating hours in the period. Total program = 6552/7276.

[d] Utilization: Hours the FGD system operated divided by total hours in the period. Total program = 6552/7276.

Source: Ref. 45.

11.1.4 Water, Chemical, and Energy Requirements

Operation of the CT-121 FGD process requires inputs of electricity, process water, cooling water, limestone, and labor. The resource requirements estimated for a 200-MW unit are given in Table 11.3. The estimated electricity requirement for the CT-121 FGD system can derate the power plant by 1.7%. Rates of water and limestone requirements include 3.5 gal process water, 5.6 gal cooling water, and 2.9 lb limestone/lb sulfur removed.

The limestone-input requirement of CT-121 is directly affected by limestone utilization. Test results from the Scholz facility indicate that the CT-121 process can achieve greater than 98% limestone utilization.[45] Limestone utilization decreases with increasing pH of the absorbent. In addition to pH, limestone utilization is affected by the residence time of liquid in the reactor. An extended residence time (greater than one hour) is needed to achieve a high limestone utilization and thus a low limestone input.

11.2 ENVIRONMENTAL IMPLICATIONS

Results of pilot and prototype tests reveal that the system can achieve 85-95% SO_2 removal efficiency and 90% particulate removal efficiency. It has no NO_x removal capability, however, nor is there continued discharge of aqueous effluent from the process. The only water outputs under normal operating conditions are cooling water and the moisture included in by-product gypsum. However, occasionally purging the stream from the absorbent may be

Table 11.3 Annual Operating Requirements of a Chiyoda-121 FGD System[a]

Requirement	Scope
Electricity (10^6 kWh)	20.91
Process water (10^6 gal)	87
Cooling water (10^6 gal)	140
Limestone (ton)	36,000
Operating labor (men/shift)	2

[a]Design basis: 200 MW, 70% load factor, 3%-sulfur coal, and 90% SO_2 removal efficiency.

Source: Ref. 46.

necessary to ensure the quality of the by-product gypsum. This could depend on the quality of input limestone and fly ash.

Based on the results of a pilot test (see Table 11.4), by-product gypsum from CT-121 is of high purity and has good dewatering and handling characteristics. It can be marketed for manufacturing wallboard, portland cement, or fertilizer. Chiyoda has tested the by-product gypsum from CT-121 for use in wallboard and portland cement production. U.S. Gypsum and National Gypsum both have made successful production runs with gypsum produced by the CT-121 prototype system -- each run used more than 100 ton of the gypsum.

If market conditions do not permit its sale, gypsum from CT-121 can be disposed of by stacking, which minimizes land requirements and ponding cost. Water quality impacts of gypsum disposal is minimal because of chemical stability.

11.3 ECONOMIC MARKET FACTORS

11.3.1 Capital and Operating Costs

Table 11.5 gives the capital and operating costs for a 60-MW CT-121 FGD system, estimated in 1977 values.[42] Unit capital investment was estimated at \$31.2/kW; annualized cost, at 3.29 mill/kWh. Details of cost estimates were not given. Considering the status of process development, these estimates are relatively preliminary and should be treated as such. The same paper also presents costs for the conventional limestone scrubbing process. The CT-121 system was estimated to have a capital cost half that of limestone scrubbing and an operating cost 1-2 mill/kWh less than that of the conventional process. Chiyoda cites the following reasons that lead to lower costs: the amount of process equipment, a low land requirement, and highly efficient utilization of energy and alkali.[48]

Table 11.4 Analysis of Gypsum from Chiyoda-121 FGD Process

Characteristic	Measured Value
Particle size	
Average Stokes' diameter (m)	60
Chemical composition (% by wt)	
CaO	32.14
SO_3 (sulfate)	45.59
as $CaSO_4 \cdot 2H_2O$	98.02
$CaSO_3$ (sulfite)	Undetectable
CO_2	0.18
as $CaCO_3$	0.41
pH factor	6.8
Free water (% by wt on wet basis)	13.5
Mortar strength (psi)	
Tensile	135
Compression	640
Bending	384

Source: Ref. 46.

11.3.2 Potential Applications

The CT-121 FGD process may be applied to desulfurize flue gas from fossil-fueled boilers as well as waste gas from sources such as smelters and sulfuric-acid plants. The process produces gypsum, which can be sold as a by-product or disposed of if a market is not available.

Process Advantages

- A simple process, with low investment and operating costs.
- Potential freedom from scaling problems.
- Apparently low energy consumption.
- Use of relatively cheap limestone as absorbent.

Process Disadvantages

- Limited operating experience.
- A wet process, requiring scrubbed gas reheat.

Table 11.5 Capital and Operating Costs for Chiyoda-121 and Limestone Scrubbing FGD Systems[a]

Expenditure	Chiyoda-121 Scrubbing	Limestone Scrubbing	Limestone Scrubbing Followed by Oxidation
Capital cost ($)	1,870,000	3,800,000	4,500,000
Annual operating cost ($/yr)			
Electricity, at $0.01/kWh	85,148	89,090	110,377
Process water, at $0.1/1000 gal	4,417	4,859	4,417
Cooling water, at $0.05/1000 gal	2,103	1,000	2,103
Limestone,[b] at $10/ton	106,271	148,779	116,898
Fuel oil, at $70/kL	417,222	417,222	417,222
Low-pressure steam,[c] at $5/ton	11,431	11,431	11,431
Labor, at $15,000/man-year	120,000	120,000	120,000
Maintenance, at 3% of capital cost	56,100	114,000	135,000
Subtotal	802,692	906,381	917,448
Capital charge (17.5% of capital cost)	327,250	665,000	787,500
Overhead (10% of direct cost)	80,269	90,638	91,745
Total	1,210,211	1,662,019	1,796,693
Capital cost per output unit ($/kW)	31.2	63.3	75
Annualized cost (mill/kWh)	3.29	4.52	4.88

[a] Design bases: 60 MW, 70% load factor; 3%-sulfur coal; 90% SO_2 removal efficiency; 28°F reheat temperature.

[b] 90% through 325 mesh, 100% pure.

[c] For atomizing reheater fuel.

Source: Ref. 42.

11.4 PRESENT STAGE OF DEVELOPMENT

A research and development program, conducted in Japan by Chiyoda Corp., included operation of a 650-scfm pilot plant. The pilot plant has been in operation since 1975 and has been operated at both high- and low-sulfur fuel conditions. Flue gases to the pilot plant were generated by burning heavy fuel oils. The major parameters tested in pilot plant include: SO_2 removal, particulate removal, limestone utilization, operability, scaling, gypsum by-product, and sparge design.

The 23-MW prototype demonstration plant in Sneads, Fla., is supported by EPRI; it seeks to corroborate and demonstrate the performance, reliability, operability, and the cost and energy effectiveness of this advanced FGD process. A recent report based on results from nine-month testing indicates an SO_2 removal efficiency ranging from 85% to 90%, a 98% limestone utilization (compared to 70-90% for conventional scrubbers), and a stable SO_2 removal efficiency despite of high concentrations of chloride (up to 6000 ppm) or of fly ash. It also indicates that the CT-121 process is flexible and capable of continuous, reliable, and efficient operation -- and should represent a potentially attractive alternative to other currently available FGD technologies. Some prototype test results have been presented in Sec. 11.1 of this report. Additional results are available in Ref. 45.

12 MAGNESIA WET SCRUBBING

SUMMARY

The magnesia-slurry scrubbing technology is a regenerable, recoverable FGD process utilizing magnesium hydroxide to absorb SO_2 in a wet scrubber. The aqueous slurry of magnesium sulfite, formed in the scrubber, is dried and calcined to regenerate magnesium oxide. The magnesium oxide is hydrolized and reused in the absorber. The SO_2-rich stream is sent for sulfur recovery. The system includes relatively complicated chemical processes; however, the absorbent can be regenerated offsite to reduce operating complexity and costs. The MgO process has a 90% or greater SO_2 removal efficiency. It is relatively energy-intensive, requiring a great deal of fuel for magnesium-oxide regeneration. The magnesia-slurry FGD process is being commercially demonstrated in the U.S. on both oil-fired and coal-fired boilers. Low system availability has been a problem. Future development should also focus on improvement of system reliability and use of coal for absorbent regeneration. The latter would reduce system dependence on the nation's limited supplies of oil and gas.

12.1 TECHNICAL ASPECTS

12.1.1 Process Description

The magnesia-slurry scrubbing process utilizes a magnesia-oxide solution for SO_2 absorption and thermal regeneration of sorbent. Figure 12.1 is a simplified flow diagram of the MgO process. The entire process, excluding sulfur recovery, can be divided into three sections: prescrubber, absorber, and regenerater.

Flue gas from the air preheater of the power plant first enters a prescrubber. A fly ash slurry is injected into the prescrubber to remove particulates and chlorides. The flue gas next enters an SO_2 absorber for treatment by the magnesia slurry. Flue gas exiting the absorber is heated in a reheater and compressed in an induced-draft fan prior to entering the powerplant stack for discharge to the atmosphere.

Absorption of SO_2 by the magnesia slurry produces magnesium-sulfite crystals. A portion of the recirculating SO_2-absorber scrubbing liquor is continuously removed and directed to the regeneration facilities. Here, magnesium-sulfite crystals are separated from the liquor and thermally dried and dehydrated in a dryer. The $MgSO_3/MgSO_4$ solids are then sent to a calciner, where $MgSO_3$ is decomposed (at 1600°F) and $MgSO_4$ is reduced using carbon as the reducing agent. It is in the calciner that MgO is regenerated.

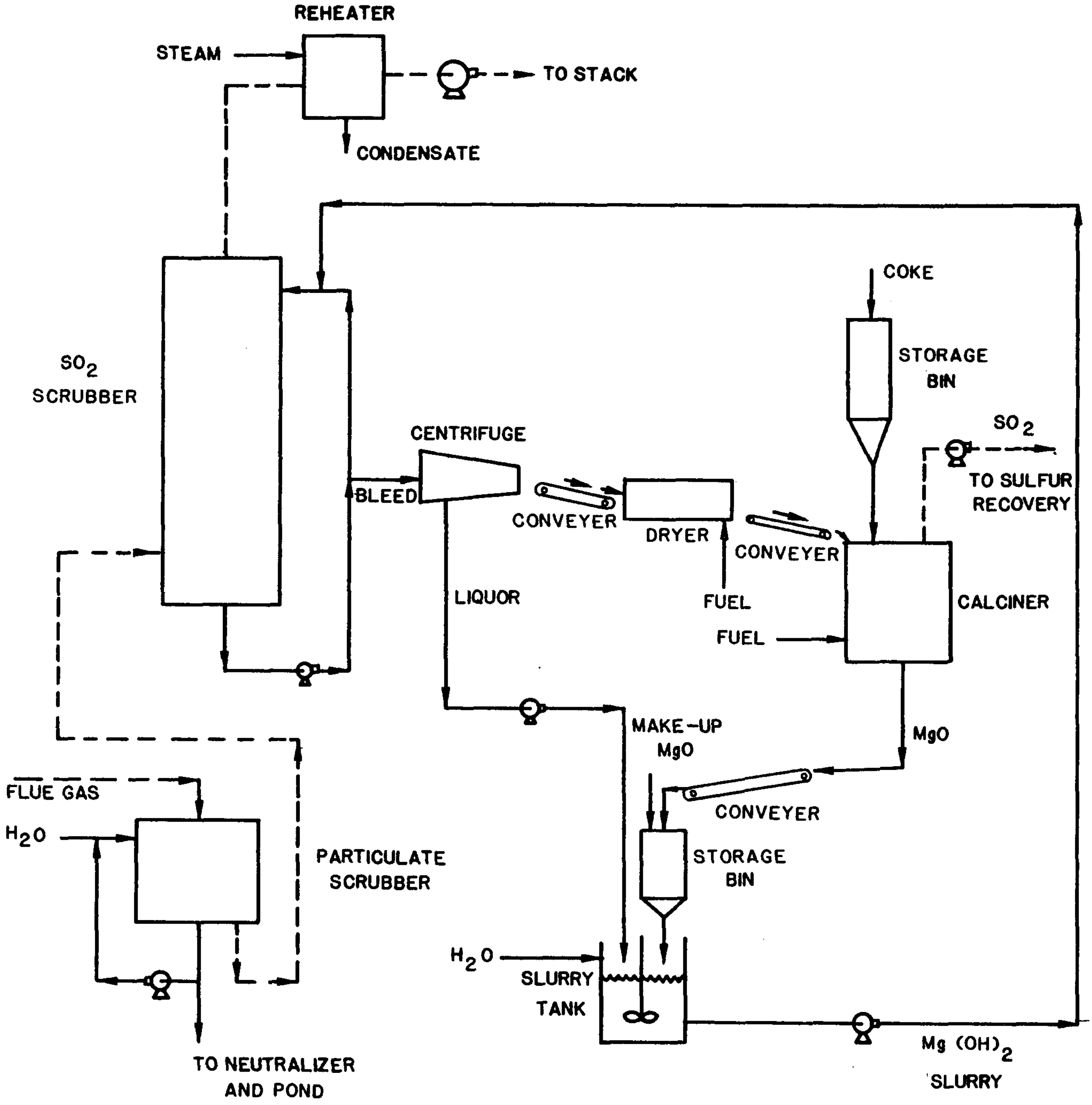


Fig. 12.1 Flow Diagram of Magnesia Wet-Scrubbing FGD Process (Source: Ref. 2)

After dust removal, the SO_2-rich gas from the calciner is piped to either a sulfur or sulfuric-acid production unit. The regenerated MgO is slaked in a slurry tank with makeup MgO to form $Mg(OH)_2$, which is recycled to the SO_2 absorber.

The material balance of a magnesia-slurry FGD system operating on a coal-fired utility boiler is shown in Fig. 12.2. System design was based on a

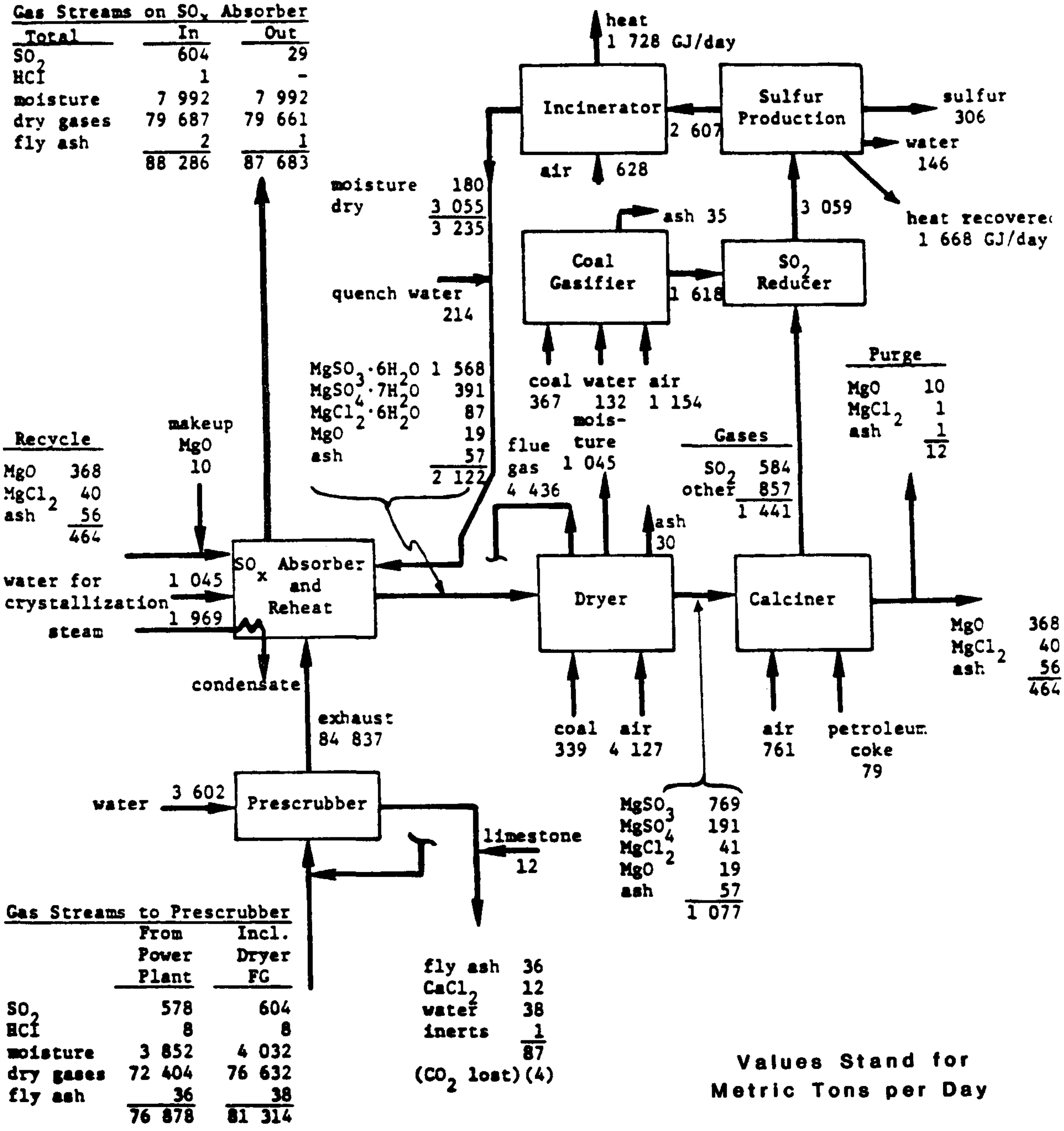


Fig. 12.2 Material Balance of a Magnesia Scrubber (Based on 95% Removal of SO_2, 95% Utilization of Mgo, and Oxidation of 17.5% Sulfate) (Source: Ref. 2)

modern 800-MW power plant using Illinois No. 6 coal, 95% SO_2 removal, 17.5% oxidation to sulfate, and 95% MgO utilization.[2] Included in the flow diagram is an SO_2 reducer, producing elemental sulfur from the SO_2 stream by using low-Btu gas. Table 12.1 lists major material flows for this facility.

Detailed descriptions of the MgO FGD process are available in the Refs. 49 through 51.

12.1.2 Process Chemistry

There are three series of reactions that occur in the magnesia-slurry scrubbing and regenerating process. The first series of reactions occurring in the scrubber includes SO_2 absorption, bisulfite neutralization, and magnesium-sulfite oxidation.

Table 12.1 Material Flow In a Magnesia System[a]

Material	Value
SO_x removal (%)	95
NO_x removal (%)	0
Input (metric t/d)	
Flue gas	81,314
Water	3,948
Fuel (coal & coke)	785
Limestone	12
MgO	10
Output (metric t/d)	
Sulfur	306
Slurry waste	87
Dry solid	77[b]

[a]For 800-MW power plant burning Illinois No. 6 coal.

[b]Including 10 ton/d MgO purge.

Source: Ref. 2.

$$Mg(OH)_2 + SO_2 \rightarrow MgSO_3 + H_2O \qquad (12.1)$$

$$MgSO_3 + H_2O + SO_2 \rightarrow Mg(HSO_3)_2 \qquad (12.2)$$

$$Mg(HSO_3)_2 + Mg(OH)_2 + 4H_2O \rightarrow 2MgSO_3 \cdot 3H_2O \qquad (12.3)$$

$$2MgSO_3 + O_2 \rightarrow 2MgSO_4 \qquad (12.4)$$

It is expected that some $MgSO_4 \cdot 7H_2O$ and perhaps $MgSO_4 \cdot 6H_2O$ will also be occluded. Magnesium sulfite along with magnesium sulfate crystals are removed from the scrubbing system and dried and dehydrated in a dryer. The chemical reactions that occur in the dryer are as follows.

$$MgSO_3 \cdot 3H_2O \rightarrow MgSO_3 + 3H_2O \qquad (12.5)$$

$$MgSO_4 \cdot 7H_2O \rightarrow MgSO_4 + 7H_2O \qquad (12.6)$$

The dry crystals are calcined in a reducing atmosphere to regenerate MgO and SO_2. The reactions occurring in the calciner are

$$MgSO_3 \rightarrow MgO + SO_2 \qquad (12.7)$$

$$MgSO_4 + 1/2C \rightarrow MgO + SO_2 + 1/2CO_2 \qquad (12.8)$$

12.1.3 Process Performance

SO_2 Removal Efficiency. Three full-scale MgO FGD units have been installed and tested in the U.S. (See subsection immediately below and Table 12.2.) Sulfur dioxide removal efficiency at all three locations were in excess of 90%. Guidelines to achieve high efficiency include the following.[35]

1. High-efficiency particulate removal should precede the SO_2 absorption.
2. A prescrubber should be used to remove any remaining particulates and most of the chlorides and SO_3.
3. Venturi absorbers should be utilized and should typically operate at a pressure drop of 25 cm (10 in.) of water or greater; or turbulent contact absorbers (TCA) should be used, operating at approximately 20 cm (8 in.) of water pressure drop and at a liquid-to-gas ratio of 5.3 to 6.6 L/normal m^3 (40 to 50 gal/100 acf).
4. The absorber superficial gas velocity should not exceed 3 m/sec (10 ft/sec).
5. The slurry pH measured at the absorber discharge should be maintained in the 6-7.5 range.

Coupled with an ESP (electrostatic precipitator), the MgO process can reduce the particulate emission by more than 99%. In the past, the MgO systems installed have not had overall performance guarantees. Now, however, certain process suppliers are willing to guarantee the entire MgO system

Table 12.2 Full-Size U.S. Demonstrations of Magnesia Systems, by Utility

Identifying Description	Boston Edison	Potomac Electric & Power	Philadelphia Electric
Station/ location	Mystic #6/ Everett, Mass.	Dickerson #3/ Dickerson, Md.	Eddystone #1A/ Eddystone, Penn.
Size (MW)	155	95	120
Boiler fuel			
Type	Fuel oil	Coal	Coal
Sulfur (%)	2.5	2	2.5
Startup date	April 1972	Sept. 1973	Sept. 1975
Status	Not currently operational	Not currently operational	Currently operational
Vendor	Chemico	Chemico	United Engineers

Source: Ref. 11.

mechanically, as well as to specify that the unit will meet applicable SO_2 emission regulations, including a 90% removal efficiency.

System Reliability. Magnesium-oxide scrubbing has received considerable attention in the past several years. Development of the process in the U.S. has been undertaken primarily by two companies: Chemico, and Babcock and Wilcox. In addition, United Engineers and Constructors is actively involved in design technology. Three full-size MgO slurry units installed on utility boilers in the U.S. are listed in Table 12.2. Operation of demonstration units at an EPA/Boston Edison unit (150-MW) and a Potomac Electric and Power Co. unit (95-MW) has been terminated. The 120-MW system at Philadelphia Electric was operational, but loss of an offsite calcining facility caused a temporary system shutdown.

The demonstration systems at Boston Edison and Potomac Electric Power were prototype units built to demonstrate the potentials of the process and to determine the major areas for improvement. These units were built on a low budget and include little redundancy. The operability* of the Boston Edison unit for the entire test run was between 13% and 81%. The unit worked best during its last four months, when operability was about 80%. The operation problems experienced at the Boston Edison unit were mainly related to the handling of magnesium-sulfite crystals. The system was designed to handle magnesium-hexahydrate crystals; instead, however, trihydrate crystals were produced in the absorber. The trihydrate crystals are much finer than hexahydrate crystals and caused dusting problems in calciners.

Operation of the magnesia-scrubbing system at Dickerson station of Potomac Electric Power Co. was on an intermittent basis between Sept. 1973 and Aug. 1975, and this system was also impeded by problems. The operability was between 27% and 68%. Mechanical and material failures due to erosion and corrosion were the main reasons the system was shut down. MgO slaking and handling were difficult because of differences between the regenerated and virgin MgO fed to the system.

The MgO facility at Philadelphia Electric, scheduled for startup in 1975, had to be temporarily shut down in Jan. 1976 because of the shutdown of a sulfuric-acid plant. The operability data available are limited to the startup phase. From Oct. 2, 1975, through Dec. 31, 1975, operability of the SO_2 scrubber was only 33%. Problems similar to those experienced at the other two demonstration plants have occurred.

Readers are referred to Refs. 1, 52, 53, and 54 for further discussions of the MgO-slurry test results.

Magnesium-slurry scrubbing has been applied on a commercial scale at three locations in Japan. These units have shown an SO_2 recovery of more than 90%. A summary of the application of these Japanese installations is shown in Table 12.3.

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*Operability is defined as hours of the FGD operation divided by hours of boiler operation in a given period, expressed as a percentage.

Table 12.3 Commercially Operating Magnesia Systems in Japan

Company/ Location	Type of Plant	Process Vendor	Flue Gas SO_2 Content (ppm)	SO_2 Recovery (%)	By-Product (ton/day)	Startup Date
Onahama Smelter (Onahama, Japan)	Copper Smelter	Onahama-Tsukishima	15,000-25,000	99.5	H_2SO_4(240)	Dec. 1972
Mitsui Mining (Hibi, Japan)	H_2SO_4	Mitsui-Mining	1500-2000	90	H_2SO_4(18)	Oct. 1971
Idemitsu Kosan (Chiba, Japan)	Claus unit and boiler	Chemico-Mitsui	-	95	Sulfur(70)	1974

Source: Ref. 11.

Sulfur Recovery. The calcination facility in the MgO slurry FGD system produces an SO_2 gas stream suitable for recovery of sulfur. The SO_2 stream from the process can be expected to be relatively low in SO_2 compared to those from such other recoverable processes as Wellman-Lord, Bergbau-Forchung, and absorption/steam.[55] In addition, it contains magnesium-oxide dust that must be removed before being sent to sulfur-recovery units. Although SO_2 can be converted into elemental sulfur or sulfuric acid, conversion to sulfuric acid seems preferable because the SO_2 stream from the MgO process contains 5 to 10% O_2.[56] Oxygen is beneficial to the production of sulfuric acid, but it increases the quantity of reducing gas required to produce elemental sulfur.

12.1.4 Water, Chemical, and Energy Requirements

Water. Fresh water is added to an MgO slurry; it serves many purposes, including:

1. Saturation of flue gas.
2. Particulate/chloride removal.
3. MgO-slurry preparation.
4. Sulfur recovery.

Under the normal operating conditions, the system loses water in four ways.

1. Evaporation loss.
2. Prescrubber blowdown.
3. Loss with solid drying.
4. SO_2 conversion to H_2SO_4 or elemental sulfur.

The estimated makeup-water requirement for an MgO system operating on a power plant is shown in Table 12.4. These data are for a 500-MW power plant operating at an efficiency of 37%, burning 3.5%-sulfur coal, and achieving a 90% SO_2 removal efficiency.[35] Total water consumption for the MgO system was estimated to be 580 gal/min, or 1.158 gal/min per MW. More than 80% of this requirement is used to make up the evaporation loss. An additional 130 gal/min would be required if sulfuric acid is produced; 3 gal/min if elemental sulfur is produced. These requirements could vary depending on the selection of specific sulfur-recovery systems.

Chemicals. Major chemical requirements of a 500-MW MgO FGD system operating on a boiler burning 3.5%-sulfur coal and requiring a heat input of 12,000 Btu/lb were given as: MgO makeup, 420 lb/h; hydrated lime, 800 lb/h. It also requires $145 \cdot 10^6$ Btu/h low-Btu gas for sulfur production (not required for H_2SO_4 production), or, 0.2 L/h catalyst if H_2SO_4 is produced.[44]

Energy. A detailed analysis of energy requirements for the MgO process was conducted by Radian Corp.[34] The unit considered was based on 90% SO_2 removal from the flue gas of a 500-MW power plant burning 3.5%-sulfur coal. Particulates and chloride are removed from the flue gas prior to the SO_2 scrubber. The scrubbed gas is reheated by using indirect steam heaters to achieve a stack-exit temperature of 175°F. Oil-fired dryers and calciners are used to decompose $MgSO_3$ into solid MgO and gaseous SO_2 streams. The SO_2 from calciners is converted to sulfuric acid by using a conventional contact sulfuric-acid plant. Altogether, nine general operations in the MgO FGD

Table 12.4 Makeup Water Requirements for a Magnesia System[a]

Source of Water Loss	Makeup (gal/min)
Evaporative loss	480
Prescrubber blowdown	54
Solid drying	45
SO_2 conversion requirement	
Sulfuric acid	130
Elemental sulfur	3
Total: If sulfuric acid is produced	790
If elemental sulfur is produced	582

[a] Basis: 500-MW power plant operating at an efficiency of 37%, burning 3.5%-sulfur coal, achieving 90% SO_2 removal efficiency.

Source: Ref. 35.

process were considered. These operations and their oil, steam, and electricity requirements are listed in Table 12.5. The operation of this FGD system requires 10,170 kW of electricity, 44.5 · 10^6 Btu/h of steam, and 113.8 · 10^6 Btu/h of fuel oil. Assuming that all these electricity and steam requirements are obtained from the power plant, the MgO system would derate the power plant by 16 MW, or 3.2% of the generating capacity. The overall energy penalty for the MgO system, including electricity, steam, and fuel oil requirements, is estimated to be equivalent to 6.06% of energy input to an uncontrolled 500-MW power plant.

Magnesia scrubbing is a relatively complicated process involving high-temperature operation. As expected, it requires more energy for operation than nonregenerable systems such as lime and limestone. MgO regeneration and SO_2 production are the steps where the largest energy consumption occurs. Process and equipment developments need to be made in these areas to improve thermal efficiency and to make use of coal in place of fuel oils.

The system uses less energy than such other regenerable systems as Wellman-Lord, according to the energy-balance calculations made by Radian.[34]

Table 12.5 Energy Requirements for a Magnesia FGD System[a]

Requirement	Electricity (kW)	Fuel Oil (10^6 Btu/h)	Steam (10^6 Btu/h)
Raw material handling and preparation	180	-	-
Particulate/chloride removal	1,090	-	-
SO_2 scrubbing	910	-	-
Reheat	-	-	44.5
Fans	5,420	-	-
Slurry processing	380	-	-
Cake drying	410	54.5	-
$MgSO_3$ calcining[b]	380	59.4	-
Sulfuric-acid production	1,260	-	-
Utility and service	140	-	-
Total	10,170	113.8	44.5

[a] Basis: 500-MW power plant burning 3.5%-sulfur coal, with 90% SO_2 removal.

[b] The waste-heat boiler in $MgSO_3/MgSO_4$ calcining was estimated to produce the equivalent of 14.4 10^6 · Btu/h of steam. This steam has been credited to the heat duty for flue-gas reheater and slurry-processing operation.

Source: Ref. 34.

For a magnesia-slurry system, the energy requirement given, as a percent of the energy input to an equivalent uncontrolled power plant, ranges between 5% and 10%. These compare to energy requirements of 3-4.5% for lime/limestone processes and 12-25% for the Wellman-Lord/Allied Chemical FGD process.

12.2 ENVIRONMENTAL IMPLICATIONS

12.2.1 Gaseous Emissions

Existing MgO systems have consistently shown an SO_2 removal efficiency of 90% or greater. The enriched SO_2 stream can be processed to recover sulfur. Coupled with an electrostatic precipitator, the MgO process can reduce particulate emissions by more than 99%. This FGD process has no capability for NO_x removal.

12.2.2 Liquid Effluents

The process has two liquid effluents. The prescrubbing system blowdown may have a chloride content of 10,000 to 20,000 mg/L, 5% suspended solids, and trace amount of amounts of chemicals from the fly ash. An intermittent purge of the magnesia slurry could contain silica, ferric oxide, aluminum chloride, sulfate-calcium oxide, calcium sulfate, calcium chloride, magnesium sulfite, magnesium sulfate, and trace elements. These effluents can be treated and the purified water discharged or made available for recycling.

12.2.3 Solid Wastes

As a regenerable process, MgO produces some solid waste from purged materials. For an MgO system operating on an 800-MW coal-fired power plant, a purge of waste solids at 12 metric t/d was estimated to be required to hold the inerts to 20% in the circulating solids (see Fig. 12.2). The purged solids contain MgO (80% by wt), magnesium chloride (8.3% by wt), and coal ash (11.7% by wt). This waste appears to produce minimal disposal problems, but studies to carefully define its environmental impact are needed.

12.3 ECONOMIC AND MARKET FACTORS

12.3.1 Operating and Capital Costs

Cost evaluation for the MgO system has been conducted in several independent studies.[2,44,57] Due to process sophistication, capital and operating expenses for this system appear to be higher than for nonregenerable systems and several other regenerable systems. Table 12.6 provides cost comparisons between the MgO system and other FGD systems.

A conceptual design and preliminary cost estimate of a utility MgO FGD system are presented by TVA.[44] The system is a new 500-MW midwestern power unit burning 3.5%-sulfur coal. Electrostatic precipitators are designed

Table 12.6 Comparative Costs of Magnesia and Other FGD Systems

	For 500-MW Plant[a]		For 800-MW Plant[b]		For 500-MW Plant[c]	
FGD System	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)
Nonregenerable						
Lime	90	98	104	6.1	124	9.68
Limestone	98	4.02	109	5.9	142	10.15
Regenerable						
Sodium-sulfite-based double-alkali	101	4.19	143	7.2	147	11.02
Magnesia (to H_2SO_4)	132	5.08	150	7.5	156	11.36
Wellman-Lord (to H_2SO_4)	131	5.11	d	d	141	9.84
Wellman-Lord (to sulfur)	148	6.14	143	7.6	d	d
Citrate	143	6.44	137	7.4	d	d

[a] Midwestern power plant burning 3.5%-sulfur coal; 78.5% SO_2 removal; 30-yr operating life; 1979 dollars; by-product credits included. Source: Ref. 5.

[b] Burning 3.5%-sulfur coal; 95% SO_2 removal efficiency, except 90% on limestone; 1978 dollars; credits for by-products (sulfur and H_2SO_4) included. The Mgo system produces elemental sulfur. Source: Ref. 2.

[c] Burning 3.5%-sulfur coal; 78.5% SO_2 removal efficiency; onsite waste disposal; 1980 dollars; spare modules assumed to be allowed; by-product credits included. Source: Ref. 57.

[d] Not available.

to remove 99.5% of the particulates in the flue gas and are used ahead of the MgO FGD system. The cost for removal and disposal of the fly ash was not included.

The MgO system designed by TVA include the following features.

- HCl in the flue gas is removed (in a prescrubber), neutralized, and discarded.
- SO_2 removal was based on meeting 1.2 lb SO_2 allowable emission/10^6 Btu heat input (or 78.5% SO_2 removal).
- Indirect steam is used to reheat the scrubbed gas prior to emission through plant stack.
- $MgSO_3$ solids are dried and dehydrated in an oil-fired rotary kiln and decomposed for reuse in an oil-fired fluid-bed calciner.
- SO_2 is converted to H_2SO_4 in a single-contact acid plant. Each year a total of 108,000 tons of 100% sulfuric acid is produced.

The estimated capital investment for the MgO slurry FGD system is summarized in Table 12.7. It includes total direct investment and allowances for startup and modification, interest during construction, and working capital. Total capital investment was estimated at $65,911,000 (1979 dollars); unit capital investment, at $132/kW.

Average annual revenue requirements were estimated based on 7000 h/yr of operation; 1980 raw material, labor, and utility costs; straight-line depreciation of 3.5% of the total depreciable capital investment; 0.7% interim replacement allowance; a property tax of 1.5%; an insurance allowance of 0.5% of total depreciable capital investment; and $25/ton credit for the by-product sulfuric acid. Net average annual revenue requirements were estimated at $18,325,000 (1980 dollars), and the net equivalent unit revenue requirements at 5.08 mill/kWh. These requirements are summarized in Table 12.8. A fairly significant part of operating requirement derives from steam used for scrubbed flue-gas reheat and from fuel oil used for drying and recalcining $MgSO_3$.

12.3.2 Potential Applications

The relative complexity of this process may limit its application to utility or the larger industrial boilers or to multiple-boiler situations where multiple absorbers and a common regenerator might be installed. The magnesia process has the following advantages and disadvantages.

Advantages

- Greater than 90% SO_2 removal efficiency.
- Recoverable process producing marketable sulfuric acid or sulfur by-product.

Table 12.7 Summary of Estimated Capital Investment for a Magnesia System[a]

Capital Expenditure	Investment ($)	% of Total Direct Investment
Direct Investment		
Materials handling (conveyors, silos, bins, feeders)	704,000	2
Feed preparation (mixer, tank, agitator, pump)	303,000	0.9
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts, dampers from absorber to reheater and stack)	4,625,000	13.1
SO_2 absorption (four spray grid scrubbers including entrainment separators, tanks, agitators, and pumps)	5,447,000	15.4
Stack gas reheat (four indirect steam reheaters)	1,198,000	3.4
Chloride purge (four chloride scrubbers and entrainment separators, tanks, agitators, pumps)	5,066,000	14.3
Slurry processing (centrifuges, conveyor, tank, agitator, pumps)	1,110,000	3.1
Cake drying (dryer, conveyors, silos, fans, tank, pumps)	5,518,000	15.7
Calcination (calciner, preheater, solids cooler, waste heat boiler, conveyors, silos, fans, bins)	2,243,000	6.3
Acid production (complete contact unit for sulfuric acid production	6,000,000	17
Acid storage (storage and shipping facilities for 30-d production of sulfuric acid)	994,000	2.8
Subtotal	33,208,000	94
Services, utilities, and miscellaneous	1,992,000	5.6
Total process areas excluding pond construction	35,200,000	99.6
Incremental ash-pond cost	154,000	0.4
Total direct investment	35,354,000	100
Indirect Investment		
Engineering design and supervision	1,881,000	5.3
Architect and engineering contractor	467,000	1.3
Construction expense	4,831,000	13.7
Contractor fees	1,442,000	4.1
Total indirect investment	8,621,000	24.4
Contingency	8,795,000	24.9
Total fixed investment	52,770,000	149.3
Other Capital Charges		
Allowance for startup and modifications	5,262,000	14.9
Interest during construction	6,333,000	17.9
Total depreciable investment	64,365,000	182.1
Land	42,000	0.1
Working capital	1,504,000	4.2
TOTAL CAPITAL INVESTMENT:	65,911,000	186.4

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur in coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; 6.5 ton/h 100% H_2SO_4.

Evaluation represents project beginning mid 1977, ending mid 1980. Average cost basis for scaling, mid 1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

Source: Ref. 58.

Table 12.8 Summary of Average Annual Revenue Requirements for a Magnesia System[a]

Annual Revenue Requirement	Annual Quantity	Unit Cost ($)	Total Annual Cost ($)	% of Average Annual Revenue Requirements
Direct Costs				
Raw materials				
Magnesium oxide (ton)	1,470	300	441,000	2.48
Catalyst (L)	1,800	2.50	4,500	0.03
Hydrated lime (ton)	3,240	15	48,600	0.27
Total raw materials cost			494,100	2.78
Conversion costs				
Operating labor/supervision (man-hour)	47,500	12.50	593,800	3.34
Utilities				
Fuel oil No. 6 (gal)	6,286,000	0.40	2,514,300	14.13
Steam (10^6 Btu)	503,400	2	1,006,800	5.66
Process water (kgal)	2,359,200	0.12	283,100	1.59
Electricity (kWh)	61,752,000	0.029	1,790,800	10.07
Heat credit (10^6 Btu)	135,600	2	(271,200)	(1.52)
Maintenance				
Labor and material			2,468,600	13.88
Analyses (man-hour)	8,500	17	144,500	0.81
Total conversion costs			8,530,700	47.96
Total direct costs			9,024,800	50.74
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			3,861,900	21.71
Average cost of capital and taxes at 8.6% of total capital investment			5,668,300	31.87
Overheads				
Plant, 50% of conversion costs less utilities			1,603,500	9.01
Administrative, 10% of operating labor			59,400	0.33
Marketing, 10% of by-product sales revenue			270,000	1.52
Total indirect costs			11,463,100	64.44
TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS:			20,487,900	115.18
By-Product Sales Revenue				
100% sulfuric acid (ton)	108,000	25	(2,700,000)	(15.18)
NET AVERAGE ANNUAL REVENUE REQUIREMENTS:			18,325,000	100

	mill/kWh	$/ton coal burned	$/$10^6$ Btu heat input	$/short ton S removed
Revenue requirement per unit of input or output	5.08	11.86	0.56	512

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur fuel; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; 108,000 ton/yr 100% H_2SO_4.
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit onstream time, 7000 h/yr.
Coal burned, 1,500,100 ton/yr, 9000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short ton/yr.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $37,074,000; total depreciable investment, $68,771,000; and total capital investment, $70,293,000.
All tons shown are 2000 lb.

Source: Ref. 58.

- Regeneration of the absorbent (MgO), reducing raw material costs.
- Waste-disposal problems relatively minor.
- MgO regeneration carried out either onsite or at a site distant from the power plant, thus permitting the use of a central regenerating facility servicing several FGD units.
- Minor problems of plugging and scaling in the scrubber as a result of high circulation rates, control of slurry composition, high concentration of crystallizations nuclei, and a short residence time.

Disadvantages

- Due to process sophistication, higher capital and operating requirements than throwaway systems and several recoverable systems.
- Energy penalties relatively high.
- Relatively low reliability of past installations; many operating problems still in need of better resolution.
- A high liquid-to-gas ratio required by the scrubber.
- Extensive particulate removal needed prior to FGD processing.

12.4 PRESENT STAGE OF DEVELOPMENT

12.4.1 Current Applications

Magnesium-oxide scrubbing has had three commercial plant-scale demonstrations in the U.S. All three are retrofit units for utility boilers (one burning oil, two burning coal). This FGD process also has been applied on a commercial scale at three locations in Japan, constituting about 1% of total FGD capacity in that country. SO_2 removals of 90-99% have been demonstrated for these units. All these applications have been described in Sec. 12.1.2.

12.4.2 Planned Applications

At the time of this writing, the following four applications of the MgO FGD process are planned.[59]

Philadelphia Electric Company

- Cromby station. The utility plans to retrofit one of the two boilers (150-MW each) at Cromby with an FGD system. A contract has been awarded to United Engineers for a magnesium-oxide system. The startup date is set for 1983.

- Eddystone station unit 1B. Currently one-third of the boiler (360-MW gross) flue gas is scrubbed with MgO slurry for SO_2 removal. The installation of an MgO FGD system on the balance of flue gas from this unit will follow; operations are scheduled to commence in 1982.
- Eddystone station unit 2. United Engineers will supply a magnesium-oxide FGD system to this unit. Startup is set for December 1982.

Tennessee Valley Authority

- Jacksonville station. United Engineers has been awarded a contract to provide engineering assistance for a 600-MW equivalent FGD system at the existing 1450-MW coal-fired boiler. The system will use regenerable MgO scrubbing to remove 90% of the inlet SO_2. The system is scheduled for startup in December 1981.

12.4.3 Problems

Problem areas still exist which need to be resolved to improve MgO system reliability and process efficiency, and to reduce cost and dependence on fuels in limited supply. These areas include the following.

1. Improvement of the understanding of the basic process chemistry and kinetics including: hydrate $MgSO_3$ crystal formation; vapor-liquid mass transfer in the absorber; extent and effect of absorbent impurity buildup; and production of sulfur in calcination.
2. Solution of erosion/corrosion problems in the slurry-handling system.
3. Control or modification of the sulfite crystal type in order to reduce handling problems and energy requirements.
4. Utilization of coal rather than oil or gas in the calciner and dryer.
5. Use of coal, either directly or indirectly, for reduction of SO_2 to elemental sulfur.
6. Improved regeneration techniques for MgO, to improve sorbent activity and life.
7. Solution of MgO fines-removal problems in the dryer and calciner.

13 WELLMAN-LORD WET SCRUBBING

SUMMARY

Wellman-Lord is an aqueous process that employs a sodium-sulfite scrubbing solution to remove SO_2 from flue gas. Thermal regeneration enables the system to recover the sulfite and produce a concentrated stream of SO_2. This process has been applied commercially both in the U.S. and overseas to desulfurize flue and waste gases from oil- and coal-fired boilers, nonferrous smelters, sulfuric-acid plants, and Claus plants. This FGD processs has all of the advantages associated with sodium-sulfite-based scrubbing: a high SO_2 removal efficiency, no plugging or scaling in scrubbing, and a low liquid-to-gas ratio. It is a closed-loop operation, producing marketable end products with no large-scale solid-waste disposal problems. The regneration loop is a complicated process requiring a relatively high energy input and relatively higher capital and operating costs, therefore, than throwaway processes. Further development is needed to investigate specific process improvements; to evaluate the process performance in full-scale demonstration with coal-fired boilers; and to test the Wellman-Lord system in combination with downstream sulfur-reduction systems, specifically those using coal as a reducing agent.

13.1 TECHNICAL ASPECTS

13.1.1 Process Description

The Wellman-Lord sulfite-based scrubbing process was developed and is being marketed by Davy Powergas. It exploits the ability of a sodium-sulfite solution to absorb SO_2, thus forming a solution of sodium bisulfite, which can be thermally regenerated. A concentrated stream of SO_2 is produced, and the stream can be processed for sulfur recovery.

The process, excluding sulfur recovery, consists of four steps: (1) gas pretreatment, (2) absorption, (3) purge treatment, and (4) regeneration.

A simplified flow diagram of the Wellman-Lord FGD process shown in Fig. 13.1. Flue gas is passed through a venturi scrubber to remove the fly ash. The scrubbing liquid containing the fly ash flows into a recirculation sump, from which it is withdrawn and recycled to the venturi scrubber. A purge stream is continuously withdrawn from the sump, neutralized with slaked lime, and transported to the fly-ash-solid handling area. The flue gas proceeds into an absorber, where it contacts a countercurrent aqueous solution composed of 27% Na_2SO_3, $NaHSO_3$, and Na_2SO_4.

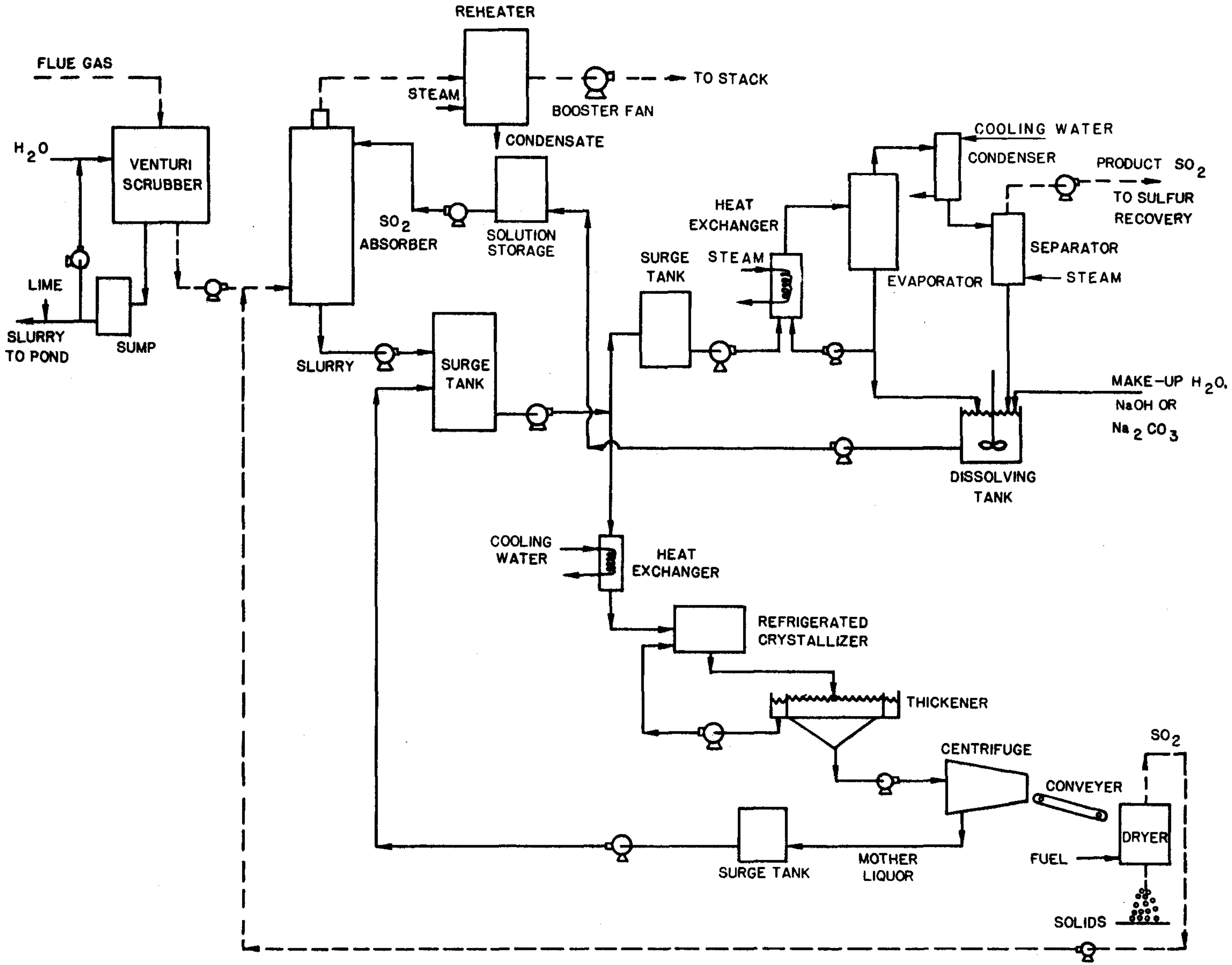

Fig. 13.1 Flow Diagram of Wellman-Lord Wet-Scrubbing Process (Source: Ref. 2)

13.1.2 Process Chemistry

The material balance for a Wellman-Lord FGD system operating on a coal-fired utility boiler is shown in Fig. 13.2. The material flows were calculated based on a modern 800-MW power plant using Illinois No. 6 coal, 95% SO_x removal, and 4% SO_3 oxidized to SO_4.[2] Included in the flow sheet is a sulfur-recovery system that reduces SO_2 to elemental sulfur by reacting it with low-Btu gas, made by gasifying coal. The waste reducing gas containing combustibles and sulfur contaminants is returned to the power plant furnace and then to scrubbing. Major material flows for this Wellman-Lord system excluding sulfur recovery are summarized in Table 13.1.

Major chemical reactions occurred in the absorber include:

$$SO_2 + SO_3^{=} + H_2O \rightarrow 2HSO_3^{-} \tag{13.1}$$

$$SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=} \tag{13.2}$$

The purified gas flows through a mist eliminator, out of the absorber. The flue gas is reheated and sent to the stack.

The spent scrubber solution from the bottom stage of the absorber is separated into two streams. One stream, about 10% of the total flow, is sent to the purge-treatment system for removal of sodium sulfate. The remainder is sent to a surge tank. From the surge tank, the absorber slurry is heated in a heat exchanger and introduced to a double-effect evaporator, where sodium bisulfite decomposes to sodium sulfite, releasing water and SO_2.

$$2HSO_3^{-} \rightarrow SO_3^{-} + SO_2 + H_2O \tag{13.3}$$

In the double-effect evaporator, a disproportionation reaction takes place at high temperature.

$$6Na + 6HSO_3^{-} \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O \tag{13.4}$$

The overhead vapors liberated from the evaporator are sent to a cooler and an SO_2 stripper. The stripped overhead vapor is cooled to reduce the water content. It is then heated, compressed, and sent to an acid or sulfur processing plant.

Condensate from the stripper is used to slurry the Na_2CO_3 crystals in a dissolving tank. Either NaOH or Na_2CO_3 is added to the dissolving tank to make up the sodium lost in the purge stream.

Sodium sulfate and thiosulfate ($Na_2SO_3O_3$) formed at the evaporator are inactive and must be removed from the system. A slipstream, therefore, is taken from the evaporator feed and sent to a purge/crystallization unit. By controlled crystallization, the sulfate precipitates in a much greater proportion than the other components. The purge stream is first cooled in a heat exchanger, followed by additional cooling in a chilled vessel. It flows to a refrigerated crystallizer, where less soluble Na_2SO_4 precipitates. The crystallizer bottoms are transferred to a thickener, from which the underflow is sent to a centrifuge, and the overflow is returned to the crystallizer.

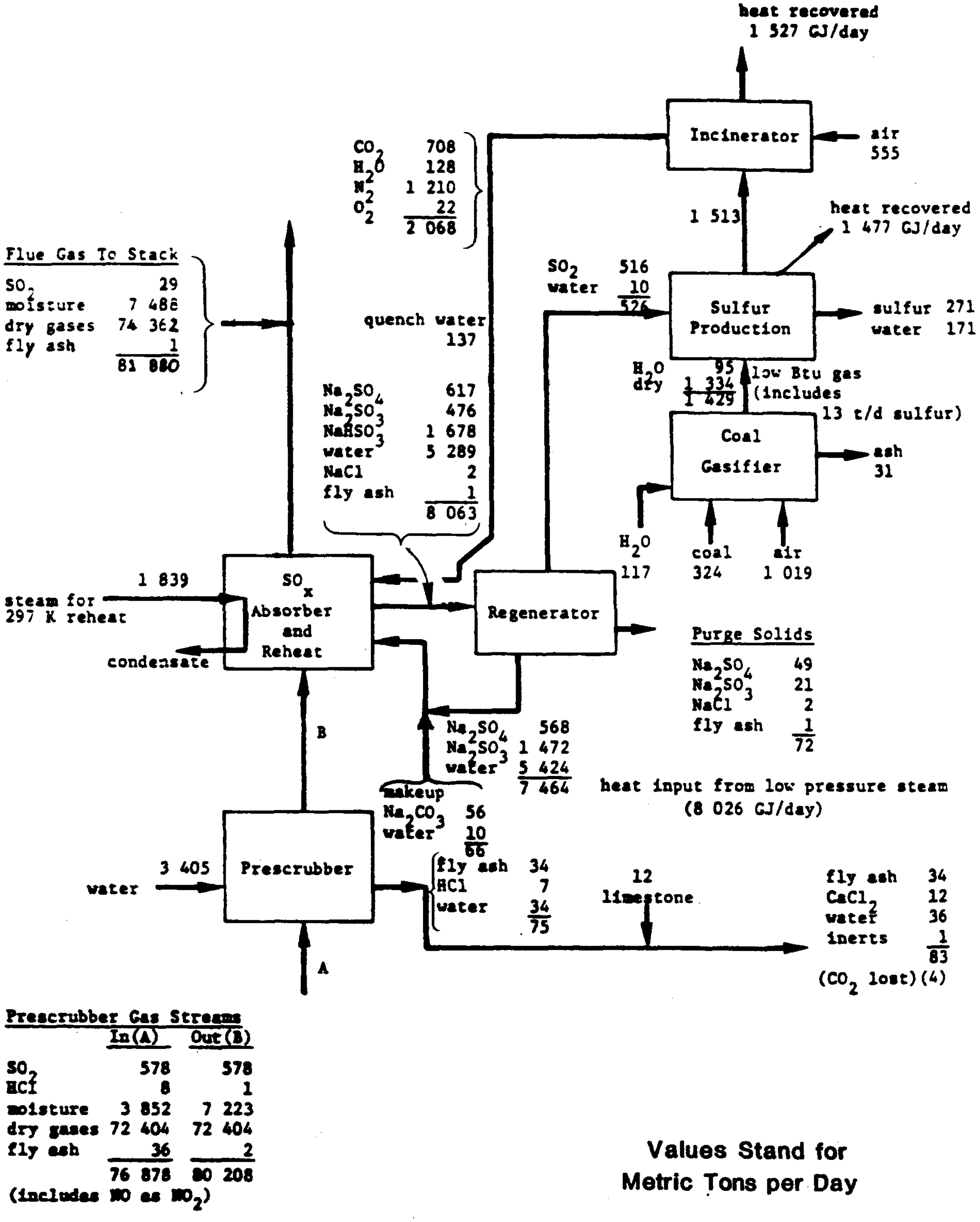

Fig. 13.2 Material Balance of a Wellman-Lord Scrubber (Based on 95% SO_x Removal, Oxidation to 4% SO_3) (Source: Ref. 2)

Wet cake from the centrifuge drops into a dryer, where sodium pyrosulfate decomposes to sodium sulfite and SO_2.

$$Na_2S_2O_5 \rightarrow Na_2SO_3 + SO_2 \qquad (13.5)$$

SO_2 vapor from the dryer is vented to the SO_2 absorber. The dried cake is transported to a storage bin, then either disposed of or sold.

Mother liquor from the centrifuge flows to a purge tank and is returned to the absorber product surge tank. Normally about 50% of the Na_2SO_4 formed during absorption and regeneration is removed in the purge treatment system.

Table 13.1 Material Flows In a Wellman-Lord System[a]

Material	Value
SO_x removal (%)	95
NO_x removal (%)	0
Input (metric t/d)	
Flue gas	76,878
Water	3,415
Low-pressure steam	8,026
Na_2CO_3	56
Limestone	12
Output (metric t/d)	
Scrubbed flue gas	81,880
Sulfur	271
Prescrubber slurry	83
Purge solids	72

[a] For 800-MW power plant burning Illinois No. 6 coal.

Source: Ref. 2.

13.1.3 Process Performance

SO_2 Removal Efficiency. The Wellman-Lord system can achieve 90-95% removal of SO_2 and may be able to achieve as high as 99% efficiency in one circumstance (high SO_2 concentration at inlet) at the expense of increased gas pressure drop and SO_3 oxidation. Since the evaporator system must be maintained free of solids, the incoming flue gas must be very low in particulates. The Wellman-Lord system, combined with high-efficiency particulate-removal systems, can remove 99% of the fly ash.

The technology can produce an off-gas containing high concentrations (90% or greater) of SO_2. The off-gas may be utilized in as a feed for conversion and recovery of sulfur by-products. At least three product options are available for conversion of gaseous SO_2 to useful chemicals: (1) liquified SO_2, (2) sulfuric acid, and (3) elemental sulfur.

Currently, the Wellman-Lord elemental sulfur-producing system employs an Allied Chemical sulfur-production unit, which uses methane as the reductant. The use of reducing gas from coal gasification in the Allied Process is still in the relatively early developmental stage. Abundant coal might be used directly as a reducing agent for producing elemental sulfur from gaseous SO_2 in conversion processes such as Foster Wheeler's RESOX. The RESOX process will be discussed in Sec. 14 of this report.

System Reliability. Considerable operating experience has been obtained with oil-fired boilers, nonferrous smelters, sulfuric-acid plants, and Claus plants. Tables 13.2 and 13.3 list commercial Wellman-Lord facilities in the U.S. and overseas, respectively.

Table 13.2 Wellman-Lord Plant Installations in the United States

Company/Location	Feed Gas Origin	Amount of Gas Treated (scfm)	Disposition of Recovered SO_2
<u>Units On Stream</u>			
Olin Corp./Paulsboro, N.J. (now shut down)	Sulfuric-acid plant	45,000	Recycle to acid plant
SOCAL/El Segundo, Calif.	Claus plant	30,000	Recycle to Claus plant
Allied Chem./Calumet, Ill.	Sulfuric-acid plant	30,000	Recycle to acid plant
Olin Corp./Curtis Bay, Md.	Sulfuric-acid plant	78,000	Recycle to acid plant
SOCAL/Richmond, Calif.	Claus plant	30,000	Recycle to Claus plant
SOCAL/Richmond, Calif.	Claus plant	30,000	Recycle to Claus plant
SOCAL/El Segundo, Calif.	Claus plant	30,000	Recycle to Claus plant
NIPSCO/Gary, Ind.	115-MW coal-fired power plant	310,000	Elemental sulfur plant
PSCNM/Waterflow, N.M.	230-MW coal-fired power plant	3,243,500	Elemental sulfur plant
<u>Units in Design or Construction</u>			
Getty/Delaware City, Del.	Three coke-fired boilers	500,000	Sulfuric-acid plant
PSCNM/Waterflow, N.M.	530-MW coal-fired power plant	1,363,500	Elemental sulfur
Getty/Delaware City, Del.	Three coke-fired boilers	500,000	Sulfuric-acid plant

Adapted from Ref. 60.

Table 13.3 Wellman-Lord Plant Installations Overseas

Company/Location	Feed Gas Origin	Amount of Gas Treated (scfm)	Disposition of Recovered SO_2
Units On Stream			
Japan Syn. Rubber/Chiba	Oil-fired boiler	124,000	Sulfuric-acid plant
Toa Nenryo/Kawasaki	Claus plant	41,000	Recycle to Claus plant
Chubu Electric/Nagoya	220-MW oil-fired power plant	390,000	Sulfuric-acid plant
Japan Syn. Rubber/Yokkaichi	Oil-fired boiler	280,000	Sulfuric-acid plant
Sumitomo Chem./Sodegaura	Oil-fired boiler	225,000	Sulfuric-acid plant
Kashima Oil/Kashima	Claus plant	20,000	Recycle to Claus plant
Mitsubishi Chem./Mitzushima	Oil-fired boiler	373,000	Sulfuric-acid plant
Toa Nenryo/Hatsushima	Claus plant	10,000	Recycle to Claus plant
Japan Nat. Railroad/Kawaski	200-MW oil-fired power plant	435,000	Sulfuric-acid plant
Kurashiki Rayon/Okayama	Oil-fired boiler	248,000	Sulfuric-acid plant
Fuji Film/Fujinomiya	Oil-fired boiler	89,000	Liquid SO_2
Shin Daikyowa/Yokkaichi	Oil-fired boiler	253,000	Sulfuric-acid plant
Sumitomo Chem./NiihamaE	Oil-fired boiler	91,000	Liquid SO_2
Mitsubishi Chem./Mizushima	Oil-fired boiler	390,000	Sulfuric-acid plant
Mitsubishi Chem./Kurosaki	Oil-fired boiler	230,000	Sulfuric-acid plant
Tohoku Electric/Niigata	100-MW oil-fired power plant	236,000	Sulfuric-acid plant

Source: Ref. 60.

Wellman-Lord has been in operation on two oil-fired 30-MW industrial boilers in Japan since August 1971. The availability has been close to 100%. Two larger oil-fired systems, a 220-MW utility boiler and a 125-MW equivalent industrial boiler, started up in Japan in 1973 and have both been operated successfully.[35]

Nine Wellman-Lord systems are currently in operation in the U.S. Six units are installed on sulfuric-acid or Claus sulfur-recovery plants. Few operational data from these plants are available; it is reported, however, that these six units have absorber onstream times of greater than 97%.[35]

The Wellman-Lord installation at the No. 11 unit of the D.H. Mitchell generating station, Northern Indiana Public Service Co. (NIPSCO), is the first coal-fired application of this FGD process. The SO_2 stream produced from the 115-MW Wellman-Lord unit is fed to an Allied Chemical reduction unit, where elemental sulfur is produced. The one-year demonstration test beginning on Sept. 16, 1977, resulted in 89% SO_2 removal efficiency and a 50% reliability factor (hours operated/hours called upon to operate).[60,61] The reliability record was established with no redundancy built into the FGD unit. Major causes of system outages include a malfunction of damaged booster fans for maintaining high silica levels in boilers, a damaged guillotine isolation damper, and unacceptable coal quality.

Subsequently, specific process modifications were implemented, and demonstration test programs continued for another 13 months (from Oct. 1978 through Oct. 1979). The test results during this period indicate that SO_2 removal efficiency averaged 90%; the 24-h average efficiency typically ranged from 85% to 92%, corresponding to SO_2 emission from 0.25-0.34 lb/10^6 Btu of heat input. The overall average reliability was 61%, with monthly reliability factors varying from less than 10% to 99%. During a seven-month period, month-by-month reliabilities were fairly constant, primarily in a range of 70-75%. The reasons for FGD plant interruptions were both process and mechanical in nature. They include, in the order of decreasing frequency of outage: (1) SO_2 reduction unit (due to sulfur deposition, leaks, and valve repairs on a sulfur condenser); (2) the evaporator circulation pump (for mechanical reasons); (3) the booster blower (due to mechanical failure); (4) the evaporation heater (plugging); (5) the absorber (obstruction in the process water valve).[62]

13.1.4 Water, Chemical, and Energy Requirements

Water. Process water is added to a Wellman-Lord FGD system (exclusive of sulfur recovery) for the following purposes: saturation of flue gas; removal of particulates and chloride; and preparation of the Na_2CO_3 slurry. Water leaves the system for the following reasons: evaporation, prescrubber blowdown, loss in product SO_2 stream, and loss with solids drying.

In addition to the process water, cooling water is needed to remove heat from the purge stream.

The estimated process makeup water requirement for a Wellman-Lord system operating on a power plant is given in Table 13.4. These data are for

Table 13.4 Wellman-Lord Makeup Water Requirements[a]

Source of Water Loss	Makeup Requirement (gal/min)
Process Water	
Evaporative loss	480
Prescrubber blowdown	54
Loss with solids drying	34
Loss in product SO_2 stream	5
Total	573
Cooling water blowdown	220

[a]For a 500-MW plant operating at a thermal efficiency of 37%, burning 3.5%-sulfur coal, removing SO_2 at 90% efficiency.

Source: Ref. 35.

a 500-MW power plant operating at a thermal efficiency of 37%, burning 3.5%-sulfur coal, and achieving 90% SO_2 removal efficiency.[35] The water requirement for sulfur recovery is not included. Total process water consumption for this Wellman-Lord system was estimated at 573 gal/min, or 1.146 gal/min per MW. More than 80% of this requirement is for making up the evaporative loss.

Chemicals. Wellman-Lord requires a premium chemical, either caustic soda or soda ash, to make up sodium in the scrubber solution; and lime or limestone to neutralize prescrubber blowdown prior to disposal. The rates of chemical requirements were estimated to be 0.21 lb of soda ash, and 0.08 lb of hydrated lime, per pound of sulfur removed.[44] These rates, of course, could vary depending on the actual operating and design condition of the system.

Energy. The regeneration loop of the Wellman-Lord FGD system requires a considerably high energy input. As a result, Wellman-Lord system is more energy-intensive than nonregenerable systems, e.g., lime/limestone, as well as other regenerable systems such as MgO slurry. Total energy requirements of a Wellman-Lord system including sulfur recovery, as a percent of energy input to an equivalent uncontrolled power plant, range between 12% and 25% -- compared to 3-4.5% for nonregenerable systems, and 5-10% for MgO systems.[34]

A detailed analysis of energy requirements for a Wellman-Lord FGD system was carried-out by Radian Corp.[34] The FGD system considered in that study includes the following features.

1. The venturi scrubbers for particulate/chloride removal operate at a gas-phase pressure drop of 8.5 in. of H_2O, and an overall liquid-to-gas ratio of 15 gal/1000 acf.

2. The SO_2 scrubbers operate at a pressure drop of 10 in. of H_2O, and a liquid-to-gas ratio at 3 gal/1000 acf.
3. Indirect steam reheaters are used to reheat the scrubbed gas for 36°F temperature rise.
4. The purge stream is cooled in the electricity-run refrigeration system. Na_2SO_4 solids are dried by an indirect steam/air heater.
5. SO_2 regeneration is achieved by evaporation and crystallization. Heat requirement is provided by steam.

The results of the Radian study are given in Table 13.5. For this Wellman-Lord system, requirements total 10,270 kW of electricity and 331.7 10^6 Btu/hr of steam. Assuming that all these energy requirements are obtained from the power plant, the Wellman-Lord system exclusive of sulfur recovery would derate the power plant by 47 MW, or 9.4% of the generating capacity. Of all the unit operations considered, SO_2 regeneration accounts for nearly 70% of the total energy requirement for the Wellman-Lord process.

Table 13.5 Energy Requirements for a Wellman-Lord System[a]

Energy Requirement	Electricity (kW)	Steam (10^6 Btu/h)	Equivalent Energy Input to the Boiler (10^6 Btu/h)[b]
Raw material handling & preparation	60	-	0.54
Particulate/ chloride removal	1,020	-	9.18
SO_2 scrubbing	320	-	2.88
Reheat	-	45.3	51.5
Fans	6,200	-	55.8
Purge treatment	930	15.4	25.9
SO_2 regeneration	1,670	271	323
Utilities and services	70	-	0.63
Total	10,270	331.7	469.43

[a] Basis: 500-MW power plant burning 3.5%-sulfur coal and removing 90% SO_2 exclusive of SO_2 reduction.

[b] Assuming uncontrolled plant net heat rate 9000 Btu/kWh, boiler energy efficiency 88%.

Source: Ref. 34.

13.2 ENVIRONMENTAL IMPLICATIONS

13.2.1 Gaseous Emissions

The Wellman-Lord process can achieve 90-95% removal of SO_2, and, in some circumstances, may be able to achieve more than 95% SO_2 removal efficiency at the expense of increased pressure drop and SO_3 oxidation. The specific unit operations of the Wellman-Lord FGD process necessitate an ash-free flue gas feed. The FGD system, combined with high-efficiency particulate removal systems, can remove 99% or more of the fly ash. No NO_x removal has been shown for the process.

13.2.2 Liquid Effluents

The Wellman-Lord FGD system has constant aqueous effluents in prescrubbing system blowdown and condenser cooling-water system blowdown. The prescrubbing system blowdown may contain 10,000 to 20,000 mg/L chloride ions, 5% suspended solids by wt, and trace amounts of chemicals from fly ash and scrubbing liquor. The condenser cooling-water system blowdown of Wellman-Lord would have a water quality similar to that of the power plant's cooling water blowdown. These effluents can be treated for discharge or for reuse.[35]

13.2.3 Waste Solids

The sulfite-to-sulfate oxidation in the scrubbing solution requires a purge treatment that produces purge solids containing about two-thirds Na_2SO_4 and one third Na_2SO_3. The solids can be sold or otherwise disposed of. These sodium-containing solids, though small in quantity, may require special attention when disposed of in landfills.[2,44]

About 0.3 lb of purge solids is produced for every pound of sulfur removed, but this could vary depending on the extent of sulfite oxidation in the scrubbing loop.

The higher the sulfite-oxidation rate, the more purge solids and the more makeup sodium will be required. Sulfite oxidation has been the area of major concern for the Wellman-Lord process. Its rate can be affected by many factors, including sorbent impurities, scrubbing recirculation rate, temperature, oxygen content of the flue gas, contact efficiency, pH, and solution strength. The exact mechanism of sulfite oxidation, however, is still not very well understood. Various techniques for controlling oxidation, including addition of antioxidants to the recirculation system, are still being investigated.

13.3 ECONOMIC AND MARKET FACTORS

13.3.1 Capital and Operating Costs

Capital and operating requirements for the Wellman-Lord FGD system have been estimated in several independent studies.[2,44,57] Table 13.6 lists the cost estimates for the Wellman-Lord system, combined with facilities to reduce SO_2 to elemental sulfur or to produce sulfuric acid. The estimates for other FGD systems are also provided. It indicates that the capital expenses for the Wellman-Lord FGD system are higher than nonregenerable systems and several regenerable systems. The operating expenses of the process appear to be competitive if credits for by-products are included. Some reduction in the costs of the Wellman-Lord process might be achievable through improvements in design of certain unit processes, such as purge separation and oxidation control.

A conceptual design and preliminary cost estimate of a utility Wellman-Lord/Allied Chemical FGD system are presented in detail by TVA.[44] The system is a new 500-MW midwestern power unit burning 3.5%-sulfur coal with a heat rate of 9000 Btu/kWh. Electrostatic precipitators designed to remove 99.5% of the particulates in the flue gas are used ahead of the FGD system. Cost for fly ash removal and disposal was not included in the process evaluation.

The Wellman-Lord/Allied Chemical FGD system designed by TVA includes the following features.

- HCl in the flue gas is removed in a prescrubber, then neutralized and discharged.
- SO_2 removal was based on meeting 1.2 lb SO_2 allowable emission/10^6 Btu heat input (or 78.5% SO_2 removal).
- Indirect steam is used to reheat the scrubbed gas prior to emission through plant stack.
- The product SO_2 is sent to an Allied Chemical coal/SO_2 reduction system, where 32,690 ton/yr of elemental sulfur is produced.

The estimated capital investment and annual revenue requirements for the Wellman-Lord/Allied Chemical SO_2-reduction system are given in Tables 13.7 and 13.8, respectively. Total capital investment was estimated at $75,190,000 (1979 dollars). Of this requirement, about 77% is attributable to the Wellman-Lord system, and 23% to the Allied Chemical sulfur-reduction unit. Annual revenue requirements were estimated at $21,478,400 (1980 dollars), including sulfur and sodium sulfate by-product sales revenues of $1,673,500. This results in a net equivalent unit revenue requirement of 6.14 mill/kWh.

13.3.2 Potential Applications

The Wellman-Lord FGD process has the following advantages and disadvantages.

Table 13.6 Comparative Costs of Wellman-Lord and Other FGD Systems

FGD System	For 500-MW Plant[a]		For 800-MW Plant[b]		For 500-MW Plant[c]	
	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)	Investment Requirement ($/kW)	Annual Rev. Requirement (mill/kWh)
Nonregenerable						
Lime	91	4.25	104	6.1	124	9.68
Limestone	98	4.11	109	5.9	142	10.15
Regenerable						
Sodium-sulfite-based double-alkali	d	d	143	7.2	147	11.02
DOWA double-alkali	99	4.36	d	d	d	d
Magnesia (to H_2SO_4)	141	5.24	150	7.5	156	11.36
Wellman-Lord (to sulfur)	148	6.14	143	7.6	d	d
Wellman-Lord (to H_2SO_4)	143	5.45	d	d	141	9.84
Citrate	d	d	137	7.4	d	d

[a] Midwestern power plant burning 3.5%-sulfur coal; 78.5% SO_2 removal; 30-yr operating life; 1979 dollars; by-product credits included. Source: Ref. 44.

[b] Burning 3.5%-sulfur coal; 95% SO_2 removal efficiency except 90% on limestone; 1978 dollars. For recoverable FGD processes, by-product credits were included. The MgO system produces sulfur. Source: Ref. 2.

[c] Burning 3.5%-sulfur coal; 78.5% SO_2 removal efficiency; onsite waste disposal; spared modules assumed allowed; by-product credits included; 1980 dollars. Source: Ref. 57.

[d] Not available.

Table 13.7 Summary of Estimated Capital Investment for a Wellman-Lord/Allied Chemical FGD System[a]

Capital Expenditure	Investment ($)	% of Total Direct Investment
Direct Investment		
Materials handling (conveyors, silos, bins, shaker, tanks, pumps	1,056,000	2.7
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack, gas ducts from chloride scrubber to absorber)	4,699,000	11.9
SO_2 absorption (four absorbers and entrainment separators, tanks, pumps, filters, agitators, heat exchangers)	8,344,000	21.2
Stack gas reheat (four indirect steam reheaters)	1,179,000	3
Chloride purge (four chloride scrubbers and entrainment separators, tanks, agitators, pumps)	2,644,000	6.7
Sulfate crystallization (evaporator crystallizer, heat exchanger, pumps, agitator, tank, dryer, conveyers, centrifuge, bin, silo, feeder)	1,919,000	4.9
SO_2 regeneration (evaporators, heat exchangers, stripper, tanks, agitators, pumps, blower, condensers)	8,064,000	20.5
Sulfur production (complete coal reduction unit)	8,400,000	21.3
Sulfur storage (storage and shipping facilities for 30-day production of sulfuric)	593,000	1.5
Subtotal	36,898,000	93.7
Services, utilities, and miscellaneous	2,214,000	5.6
Total direct investment excluding pond construction	39,381,000	100
Pond construction	269,000	0.7
Total direct investment	39,381,000	100
Indirect Investment		
Engineering design and supervision	2,789,000	7
Architect and engineering contractor	692,000	1.8
Construction expense	5,286,000	13.4
Contractor fees	1,566,000	4
Total indirect investment	10,333,000	26.2
Contingency	9,620,000	24.4
Total fixed investment	59,334,000	150.7
Other Capital Charges		
Allowance for startup and modifications	5,907,000	15
Interest during construction	7,120,000	18
Total depreciable investment	72,361,000	183.7
Land	64,000	0.2
Working capital	1,765,000	4.5
TOTAL CAPITAL INVESTMENT:	74,190,000	188.4

[a] Basis: 500-MW new coal-fired power unit, 3.5%-sulfur coal; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; 4.7 ton/h elemental S.
Midwest plant location represents project beginning mid 1977, ending mid 1980.
Average cost basis for scaling, mid 1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located one mile from power plant.
Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

Source: Ref. 44.

Table 13.8 Summary of Average Annual Revenue Requirements for a Wellman-Lord/Allied Chemical FGD System[a]

Average Annual Requirement	Annual Quantity	Unit Cost ($)	Total Annual Cost ($)	% of Average Annual Revenue Requirements
Direct Costs				
Raw materials (ton)				
Soda ash	7,440	103	766,300	3.57
Coal	25,370	26.50	672,300	3.13
Sand	180	7.50	1,400	0.01
Catalyst			3,800	0.02
Hydrated lime	2,800	54	151,200	0.70
Total raw materials cost			1,595,000	7.43
Conversion costs				
Operating labor and supervision (man-hour)	46,500	12.50	581,300	2.71
Utilities				
Fuel oil No. 6 (gal)	582,580	0.40	233,000	1.08
Steam (10^6 Btu)	1,581,820	2	3,163,600	14.72
Process water (kgal)	5,219,260	0.12	626,300	2.92
Electricity (kWh)	48,230,700	0.029	1,398,700	6.51
Heat credit (10^6 Btu)	21,640	2	(43,300)	(0.20)
Maintenance				
Labor and material			2,756,700	12.83
Analyses (man-hour)	8,800	17	149,600	0.70
Total conversion costs			8,865,900	41.27
Total direct costs			10,460,900	48.70
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			4,341,700	20.71
Average cost of capital and taxes at 8.6% of total capital investment			6,380,000	29.71
Overheads				
Plant, 50% of conversion costs less utilities			1,743,800	8.12
Administrative, 10% of operating labor			58,100	0.27
Marketing, 10% of by-product sales revenue			167,400	0.78
Total indirect costs			12,691,000	59.09
TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS:			23,151,900	107.79
By-Product Sales Revenue				
Sulfur (ton)	32,690	45	(1,471,100)	(6.85)
Sodium sulfate (ton)	8,800	23	(202,400)	(0.94)
NET AVERAGE ANNUAL REVENUE REQUIREMENTS:			21,478,400	100

	mill/kWh	$/ton coal burned	$/$10^6$ Btu heat input	$/short ton S removed
Revenue requirement per unit of input or output	6.14	14.32	0.68	614

[a]Basis: 500-MW new coal-fired power unit, 3.5%-sulfur fuel; 1.2 lb $SO_2/10^6$ Btu heat input allowable emission; 32,690 ton/yr elemental S.
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7000 h/yr.
Coal burned, 1,500,100 ton/yr, 9000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short ton/yr.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $39,381,000; total depreciable investment, $72,361,000; and total capital investment, $74,190,000.
All tons shown are 2000 lb.

Source: Ref. 44.

Advantages

- No scaling in system.
- Low liquid/gas ratio in absorber.
- Simple operations, thus enhancing reliable performance.
- High SO_2 removal efficiency.
- Considerable operating experience with oil-fired boilers, acid plants, and Claus plants.
- Relatively low volume of waste products.

Disadvantages

- Disposal of sodium-containing purge solids causes environmental concerns.
- High energy demand and capital requirements.
- Fly ash must be kept out of the system.
- Requires soda makeup.
- Corrosive process environment requires expensive material of construction.
- Requires stack gas reheat.

The Wellman-Lord process has been applied to the flue gases from oil- and coal-fired boilers, and waste gases from nonferrous smelters, sulfuric-acid plants, and Claus plants. For its application to fossil-fired industrial boilers, the major concerns relate to high excess-air and space limitations. Relatively high oxygen atmosphere in the industrial boilers could result in increased sulfite oxidation. Space limitations could affect the selection of the system for retrofit. Other than these disadvantages, the Wellman-Lord process has a positive feature in providing the possibility of one regeneration section to economically serve several small boilers -- if the boilers are located at reasonably close distances from one to another.

As to the application of the Wellman-Lord process to utility systems, space constraints (limiting retrofit applications) probably constitute the only limitation.

13.4 PRESENT STAGE OF DEVELOPMENT

13.4.1 Installations

The Wellman-Lord FGD process was developed in the late 1960s by Davy Powergas, Inc. As indicated in Sec. 13.1.4, this FGD process is being applied commercially, both in the U.S. and overseas, to coal- and oil-fired boilers, nonferrous smelters, sulfuric acid plants, and Claus plants. A full-scale demonstration of the combined Wellman-Lord SO_2-removal process and Allied Chemical SO_2-reduction process has been funded by EPA and Northern Indiana

Public Service Co. (NIPSCO). The demonstration has been underway on a 115-MW coal-fired utility boiler at the Dean H. Mitchell station in Gary, Ind., since 1977. Operation of this plant continues.

In addition to the unit at NIPSCO, a Wellman-Lord FGD system has been installed at the Public Service Co. of New Mexico's San Juan station, located in Waterflow, N. M. Three Wellman-Lord units, with a combined capacity of 1240 MW, were designed to remove 90% of the SO_2 from the flue gas when firing coal ranging in sulfur content from 0.59% to 1.3% by wt (with an average of 0.8%). Allied Chemical SO_2-reduction trains were installed as part of FGD systems to recover elemental sulfur. These units are currently in operation. Few operational data are available, although available information indicates that these units have operated as designed and with little or no startup problems.[59]

Public Service Co. of New Mexico has decided to install one more Wellman-Lord FGD system, at San Juan station Unit 4. Startup for this unit is expected in June 1982.[59]

A Wellman-Lord FGD system has also been installed at Delaware City plant (180-MW) of the Delmara Power and Light Co. The boilers burn 7-8% sulfur coke, and the designed SO_2 removal efficiency is between 80% and 90%. Three Wellman-Lord FGD units started operation in May 1980.

13.4.2 Needs

Development work is needed to improve the process design and performance of the Wellman-Lord FGD process in the following areas.

(1) Basic chemistry of sulfite-to-sulfate oxidation in the absorber.

(2) Thermodynamics and equipment selection for the sodium-sulfate purge treatment.

(3) Energy usage in sulfite regeneration and SO_2 stripping.

Programs are also needed to review and evaluate the performance of existing full-scale Wellman-Lord systems.

In addition, tests are needed to combine the Wellman-Lord system with downstream sulfur-reduction systems using coal as a reductant; the latter systems include RESOX and the Allied Chemical coal process. Elemental sulfur is a desirable product because it can be more easily handled and transported than sulfuric acid or SO_2 gas. The use of coal would preclude the need for oil or gas in the conversion of SO_2 to sulfur.

14 RESOX SULFUR-RECOVERY PROCESS

SUMMARY

The RESOX process is one of several that have been developed to recover sulfur from SO_2-rich streams. RESOX uses coal as a direct reductant for converting sulfur dioxide to elemental sulfur, a product that appears to be more desirable in several respects than such other compounds as sulfuric acid and ammonium sulfate. The process involves relatively simple unit operations, and it could be one of the less expensive sulfur-recovery processes. However, the development of this process is not far enough along for definitive cost evaluation. Development of the RESOX process was initiated in the early 1970s by Foster Wheeler Corp. Present efforts by Foster Wheeler to move the process toward full commercialization include pilot tests and a prototype demonstration. Further system studies are needed to test its adaptability to various front-end FGD systems (e.g., Wellman-Lord) and the feasibility of using bituminous coal as a direct reductant -- as well as to improve SO_2 conversion efficiency, to demonstrate the usages of residual coal (e.g., as a fuel or SO_2 absorbent), and to improve the operation and design of sulfur-condensing and sulfur-purification systems.

14.1 TECHNICAL ASPECTS

14.1.1 Process Description and Chemistry

The RESOX process is intended to achieve a desired chemical reaction between sulfur dioxide and crushed coal, thereby processing the SO_2-rich effluents from recoverable FGD systems and producing elemental sulfur. No catalyst is employed in the process, and crushed coal is the only material consumed. The major process equipment consists of a reactor vessel and a sulfur condenser.

In the reactor vessel, countercurrents of SO_2-rich gas and crushed coal contact each other. A reaction front is developed, reducing the SO_2 to elemental sulfur, as summarized by the reaction

$$SO_2 + C \rightarrow S + CO_2 \qquad (14.1)$$

Reaction temperatures range from 1100°F to 1500°F. The air injected with the SO_2 gas burns some of the coal to maintain this temperature.

In the sulfur condenser, the gaseous elemental sulfur produced as a result of the above reaction is condensed, while the uncondensed gas containing various residual sulfur compounds (e.g., SO_2, H_2S, COS) is incinerated. That incinerated gas is returned upstream of the FGD system, establishing the continuous internal recycle stream shown in Fig. 14.1.

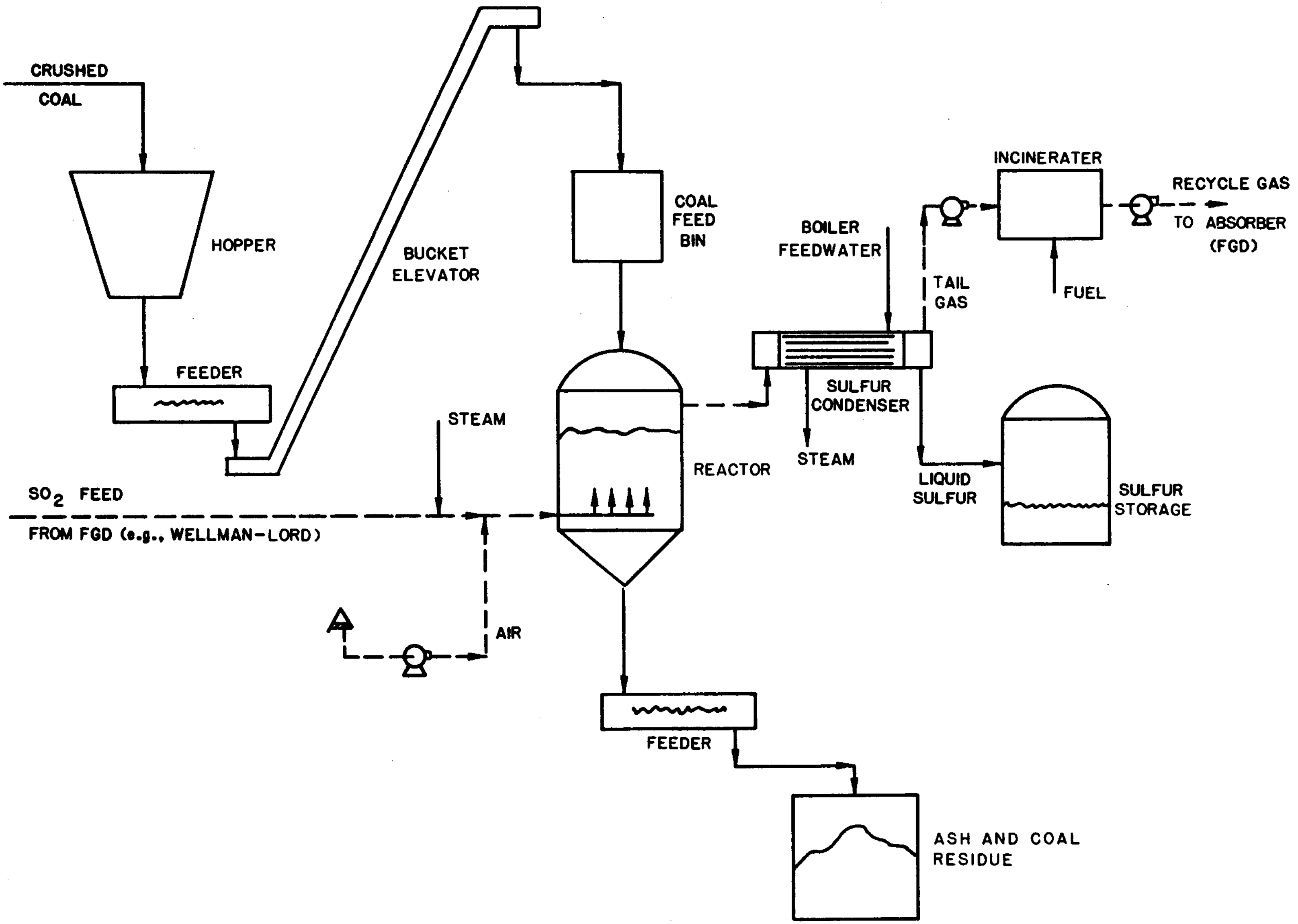

Fig. 14.1 Flow Diagram of RESOX Sulfur-Recovery Process (Applicable as Followup Stage to FGD Processes That Allow Sulfur to be Recovered) (Source: Ref. 63)

The high-purity liquid-sulfur effluent of the sulfur condenser represents a nonpolluting salable by-product. The coal-like residue at the bottom of the reactor is cooled and collected. This residue can potentially be used as fuel or SO_2 absorbent.

Details of the process chemistry, the selectivity of the various reactions as a function of temperature and SO_2 concentration, and water vapor are available in Refs. 63 and 64.

14.1.2 Process Performance

Front-End FGD Systems. RESOX requires a feed stream rich in SO_2 but low in oxygen. High SO_2 concentration is needed to maintain a high degree of reduction, while low oxygen content could prevent excessive carbon combustion. Several recoverable FGD processes, such as Wellman-Lord and Bergbau-Forschung, produce SO_2 streams of this nature and are expected to combine with RESOX without much difficulty. The Bergbau gas is quite suitable because it already has the proper H_2O/SO_2 ratio, but additional H_2O may be needed for the SO_2 stream from the Wellman-Lord process.

Conversion Efficiency. For a feed stream containing 25% or more SO_2, conversion of at least 90% of the SO_2 to sulfur is expected. Lower SO_2 content in the feed gas would result in a reduced degree of reduction. A feed stream containing 5% SO_2, for example, can be expected to reduce 75% of that SO_2. Below a 3% content of SO_2, the reduction efficiency falls off drastically.[65,66]

The sulfur produced by a limited number of prototype runs to date has a greater than 99% purity, according to a report by Foster Wheeler Development Corp.[64] Table 14.1 presents the character of the recovered sulfur product.

14.1.3 Water, Chemical, and Energy Requirements

Required inputs to the RESOX processs include crushed coal in the reactor vessel, to reduce SO_2; boiler feedwater in the sulfur condensor, as a coolant; fuel for incinerating the residual sulfur compounds; and electricity for material transport and preparation. Estimates reported at an early stage of process development indicate that 0.2 lb of coal and 0.02 kW electricity are required for every pound of SO_2 treated.[64] The estimated requirements of other resource inputs are presently not available.

Table 14.1 Sulfur Product Recovered by the RESOX Process

Character	Value
Composition (% by wt)	
Moisture	0.02
Crystalline sulfur	97.42
Amorphous sulfur	2.30
Carbon	0.24
Color	Green-yellow

Source: Ref. 64.

14.2 ENVIRONMENTAL IMPLICATIONS

The RESOX process has two waste streams: one containing noncondensing gases from the sulfur condenser, the other bearing ash and unused coal from the reactor bottom.

Noncondensing gases consist of SO_2 and some residual sulfur compounds such as H_2S and COS. These gases are not acceptable for atmospheric venting. They can be incinerated to SO_2 in an external incinerator or in a boiler furnace, and they can then be directed back to the FGD system.

The solid material that leaves the bottom of the RESOX reactor contains coal ash and partially used coal. The partially used coal appears very much the same as coal at the reactor inlet except that it has been devolatilized, demoisturized, and, to a large degree, desulfurized. The solid material, called RECOAL (a trademark of Foster Wheeler Energy Corp.), can be used either as fuel in the boiler or as an adsorbent for SO_2 (e.g., in the Bergbau-Forschung process). Significant savings in the operating costs are possible if future studies prove that RECOAL can fulfill these potentials.

14.3 ECONOMIC AND MARKET FACTORS

14.3.1 Capital and Operating Costs

The RESOX process involves only one step to convert SO_2 to sulfur. As expected, the capital cost for RESOX should be lower than for other processes such as a Claus gasification system, which involve more complicated unit operations. However, the development of the RESOX process is not far enough along for definitive cost evaluation. Furthermore, the lower capital costs may be offset by the need to use a special and more expensive type of coal, such as anthracite.

Capital cost estimated for RESOX in 1975 was $6.5/kW for a 500-MW boiler burning 4.3%-sulfur coal.[65] This is a preliminary estimate, however, and subject to change as more is learned about the process. More recent estimates are in the range of $8 to $16 per kW, considerably lower than $20 to $25 per kW estimated for a Claus gasification system.[55]

Definitive evaluation of annualized operating costs for RESOX is presently not available. One thing is certain, however: coal could make up a significant part (as much as 90%) of the total operating cost. If anthracite is required for RESOX, the coal expenses could offset the low capital requirements because of the higher cost of anthracite. Annualized cost for RESOX will much depend on whether RESOX can be adapted to bituminous coal and what the capital cost turns out to be.

14.3.2 Process Evaluation

The RESOX process is a relatively simple system designed to produce elemental sulfur from SO_2-rich streams. Several front-end FGD processes yield a product stream rich in SO_2 but low in oxygen and contaminants.

For example, processes such as Wellman-Lord, Bergbau-Forschung, and absorption/steam stripping are expected to combine easily with RESOX to produce sulfur. The RESOX process has the following advantages and disadvantages.

Advantages

- Direct reduction of SO_2, simplifying process design and operation.
- Production of elemental sulfur, which is considered a practical, economic, and environmentally desirable end product.
- Abundance and availability of coal.
- Relatively low capital and operating requirements.
- Relatively harmless enviornmental effects.
- The flexibility of producing H_2S product gas rather than a sulfur-rich stream.

Disadvantages

- The requirement of an input stream rich in SO2 but low in oxygen and containinent and the resulting possibility of excluding some FGD front-end systems.
- Relatively limited operating experience.
- Use of caking-type coals a possible problem.
- Utilization of reactor-bottom char not well defined.
- Operating problems in sulfur condensation and purification caused by the dust and other impurities in the sulfur vapor evolved from the reduction vessel.

14.4 PRESENT STAGE OF DEVELOPMENT

14.4.1 Programs

The RESOX process, conceived by Foster Wheeler Energy Corp. in the early 1970s, is being tested in pilot and prototype facilities. The history of the development of the RESOX process is highlighted by a series of important events listed below.[67]

1972 - Foster Wheeler (FW) embarked on RESOX bench-scale feasibility studies.

1973 - FW concluded RESOX pilot test work in support of a design effort by Gulf Power Co.

1975 - FW began the chemically active fluidized bed (CAFB) project, including RESOX.

1976 - Operational test runs conducted at Gulf Power Co.

1977 - Electric Power Research Institute (EPRI) and Unvelt-bundesamt (UBA) funded Lünen project, including RESOX.

Past Programs

Pilot study. Following bench-scale feasibility studies in 1972, a pilot study was initiated to collect the data necessary to design, build, and operate the Gulf prototype plant. The pilot plant was constructed at FW's Livingston, N.J., facility and operated for an extended period of time. The unit was operated with anthracite coal as the reductant. It was concluded that temperature of operation, gas residence time, dry inlet concentration of SO_2, and molar ratio of H_2O to SO_2 in the feed gas to RESOX are the primary variables affecting the process chemistry. The percentage of SO_2 conversion was shown to be dependent on the ratio of H_2O to SO_2 in the feed gas, with all other parameters held constant. Furthermore, selectivity to sulfur was found to be inversely proportional to the percentage of SO_2 converted. The pilot study also established the production of H_2S as a function of the H_2O/SO_2 molar ratio and the RESOX potential as an SO_2-to-H_2S generator.

The Gulf Project. The Gulf Power demonstration unit was erected in Chattahoochee, Fla. It treats a slipstream of flue gas from Unit 2 of the Scholz steam plant. The gas stream, equivalent to about half of the unit's maximum rating of 47.5 MW, is treated in a Bergbau-Forschung SO_2-adsorption system. The RESOX SO_2-reduction system was integrated with the FGD dry-scrubbing front-end system to produce elemental sulfur. Critical mechanical problems associated with the FGD system adversely affected RESOX operability. The RESOX subsystem was operated in an integrated mode for more than 130 h, with initial results characterized as "encouraging." The main problem with RESOX was the carryover of fine dust from the regenerator in the FGD system through the reactor to the sulfur condenser. The contaminants eventually fouled the condenser and caused premature shutdown. As a result, the Gulf demonstration unit is considered only a limited success.

On-Going Programs

The Lünen Project. A 42-MW RESOX prototype facility has been constructed and is being operated along with a Bergbau-Forschung FGD system at Lünen, the Federal Republic of Germany. The program is sponsored by EPRI (Electric Power Research Institute) and UBA (Umweltbundesamt), West Germany's environmental agency. Foster Wheeler is the prime U.S. contractor. A consortium consisting of FW, Bergbau-Forschung GmbH, STEAG A.G., and Deutsche Babcock A.G. has been formed to conduct this and related projects at the Lunen site. Construction of the RESOX unit was completed in April 1978, and the plant started up in May and June of the same year. Available results from early testing indicate that: 1) significant improvement has been made at the prototype stage, through design changes; 2) no fundamental process deficiencies were identified; 3) SO_2 conversion and outlet gas compositions are encouraging; and 4) the sulfur produced was of high purity.[64] These results are of a preliminary nature and do not warrant drawing hard conclusions regarding the ultimate commercial feasibility of RESOX.

The CAFB Project. CAFB refers to the chemically active fluidized-bed process. Heavy oils or coals are gasified in a fluid-bed reactor, in the presence of limestone and under substoichiometric conditions. Calcium sulfide is formed and then regenerated at higher temperatures to form lime and an SO_2 off-gas of low concentration. This off-gas is piped to a RESOX system for reduction to elemental sulfur.

The CAFB project, funded by the Central Power and Light Co. and the U.S. EPA, is located in San Benito, Texas. The 22-MW RESOX unit has been installed and will be tested when the front-end CAFB units are operable. The application of RESOX to CAFB produces process characteristics that differ significantly from those at the Bergbau-Forschung application: the gas stream from the CAFB is much lower in SO_2 concentration, higher in temperature, and more contaiminated. Also, RESOX is much more closely coupled with the front-end CAFB process than with Bergbau-Forschung, and this will present challenges in process control and stability.

Others. In addition to these projects, Foster Wheeler has done joint development work on RESOX with TVA and Stern Rodgers, Inc. Results from these program are presently not available.

14.4.2 Needs

For the RESOX sulfur-recovery process, it is suggested that the test program in Lünen, West Germany, should continue. Data from the CAFB project should be evaluated when available. In addition, studies should be conducted to resolve operational uncertainties and to improve process efficiency.[66] Specific studies are needed in the following areas.

Adaptability to Various Front-end Systems. RESOX has been tested in prototype demonstration units along with Bergbau-Forschung FGD systems, but tests are also needed with other FGD processes that produce desirable SO_2 streams (e.g., Wellman-Lord). The SO_2-containing off-gas from some recoverable FGD processes, such as magnesia scrubbing, seems to be less desirable due mainly to a high O_2 content. Tests are needed, however, to investigate the oxygen limit in the feed gas to RESOX.

Use of Bituminous Coal. RESOX systems have been successfully tested with anthracite as the SO_2 reductant. The process would be much more attractive economically, however, if bituminous coal could be used.

Temperature Control. Variation in oxygen content of the inlet gas stream has caused problems in the temperature control of the reactor.

Energy Efficiency. The coal is only about half consumed in the reduction step. It is proposed to use the residue either as a fuel or as an absorbent for SO_2. Demonstration of these usages is needed.

Sulfur Condensation and Purification. Dust and other impurities in the sulfur vapor from the reduction vessel may cause problems in sulfur condensation and purification.

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