MASTER

**ENVIRONMENTAL CONTROL IMPLICATIONS
OF GENERATING ELECTRIC POWER FROM COAL**

1977 TECHNOLOGY STATUS REPORT

APPENDIX F

**FLUE GAS DESULFURIZATION
IN THE UNITED STATES - 1977**

**PREPARED BY
TENNESSEE VALLEY AUTHORITY**

**ENVIRONMENTAL CONTROL-
COAL UTILIZATION PROGRAM**

ARGONNE NATIONAL LABORATORY

**PREPARED FOR
DIVISION OF ENVIRONMENTAL CONTROL TECHNOLOGY
ASSISTANT SECRETARY FOR ENVIRONMENT
UNITED STATES DEPARTMENT OF ENERGY**

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ARGONNE NATIONAL LABORATORY
9700 South Cass Avenue
Argonne, Illinois 60439

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Appendix F
Flue Gas Desulfurization
in the United States - 1977

Prepared by
Tennessee Valley Authority
Muscle Shoals, Alabama
for
Argonne National Laboratory

December 1977

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FOREWORD

"Environmental Control Implications of Generating Electric Power from Coal" is a continuing Argonne National Laboratory program sponsored by the Division of Environmental Control Technology of the U.S. Department of Energy. This 1977 Technology Status Report is the third in a series of reports issued as part of the program and represents efforts of Argonne and several sub-contractors. The primary emphasis is on characterizing and evaluating recent developments in available and near-term control technologies through detailed engineering and cost analyses. The report also includes an assessment of the effects of recent regulatory developments and comparative evaluations of several possible control technology combinations.

The main volume of the report is supplemented by seven appendices consisting of subcontractor reports that deal with particular control technologies. Each appendix is a separate volume, as indicated below, and is designed to be understandable without reference to other portions of the report.

This appendix (F) was prepared by F. M. Kennedy and S. V. Tomlinson of the Tennessee Valley Authority, Muscle Shoals, Alabama. The work was carried

Appendix, Title	Subcontractor
A* Coal Preparation and Cleaning Assessment Study (Part 1), and Appendix (Part 2)	Bechtel Corp.
B Assessment of Status of Technology for Solvent Refining of Coal	Air Products and Chemicals, Inc.
C Assessment of Low-Btu Gas Combined Cycle Power Generation	Foster Wheeler Energy Corp.
D Assessment of NO _x Control Technology for Coal Fired Utility Boilers	KVB, Inc.
E A Review of Technology for Control of Fly Ash Emissions from Coal in Electric Power Generation	Southern Research Institute
F Flue Gas Desulfurization in the United States - 1977	Tennessee Valley Authority
G State of the-Art Review for Simultaneous Removal of Nitrogen Oxides and Sulfur Oxides from Flue Gas	Tennessee Valley Authority

* Appendix A is published as two separately bound volumes. Part 1 is the main text, and Part 2 consists primarily of data generated by merging coal washability and reserve data obtained from the U.S. Bureau of Mines.

out under ERDA-TVA Interagency Agreement EY-77-A-02-4201, and will also be reported in TVA Bulletin Y-125. Management of the project was the responsibility of G. G. McGlamery of the Tennessee Valley Authority and P. S. Farber of Argonne National Laboratory.

The principal volume of the report and the various appendices are being issued independently as they are completed, in order to make the information available in a timely manner. Inquiries regarding the availability of specific volumes should be directed to Technical Information Services, Argonne National Laboratory, Argonne, Illinois 60439.

EXECUTIVE SUMMARY

Electrical generation will depend increasingly on coal and nuclear fuels for its estimated growth of 5.3%/yr. Following 1970, many power plants changed from coal to low-sulfur oil for environmental reasons; now all future and many existing plants are being directed to use coal. Nevertheless, the amount of oil used for electrical generation (but not oil's share of total fuels) will increase for several years before tapering downward. The expansion and the reversion to coal will increase the need for SO₂ control, especially by flue gas desulfurization (FGD).

Published literature on FGD for coal-fired power plants often references a nominal SO₂ emission standard of 1.2 lb SO₂/MBtu heat input to the boiler. This is the present NSPS level for units whose construction started after August 17, 1971. Actually, utilities operate by one or more of several hundred local SO₂ control levels contained in SIPs which are under the aegis of EPA. For new plants, SIPs may equal the NSPS level or be a fraction of it; for older plants, SIPs range both below and above the NSPS level. In any case, the SIP for each plant dictates the degree of effectiveness required for any SO_x control option.

Many power plants are looking to low-sulfur coal, mainly from Montana and Wyoming, as an alternative to FGD. This can be advantageous for plants with favorable locations and other conditions, but many factors (availability of coal and means of transportation, plant conversion cost, delivered coal cost, fuel cost adjustment clauses, type of boiler and ESP, plant derating, etc.) prevent it from being a universal option. Its strongest case is for older plants. There are situations, of course, where a stringent SIP will still require FGD when burning low-sulfur coal.

Coal desulfurization or "front-end treatment" varies widely in effectiveness and availability for plant use. The long-standing physical coal cleaning methods remove clay and ash but very little sulfur. Deep physical cleaning (heavy-media or other separations) can remove up to 85% of the pyritic sulfur (but little or no organic sulfur) as medium-sulfur middlings and high-sulfur tailings. Its overall advantage is limited by coal loss to tailings and by the need for FGD or other sulfur control during middlings combustion. In terms of the total desulfurization picture both classes of physical cleaning are limited to special situations, but physical coal

cleaning and FGD will have economical combinations. Chemical coal cleaning and fluidized-bed combustion have good technical potential but they, like coal conversions (gasification and/or liquefaction), will need substantial resources and time for development to operational status.

Forty FGD processes, some with vendor variations, are known to be reasonably active in the world; many of these exist at pilot-plant or smaller scale. Several dozen claimed processes are not considered active because of being abandoned, dormant, or not beyond an idea or a proposal. In most process classification systems, the whole process is categorized as throwaway versus recovery, wet versus dry, etc., but difficulties often arise from such simplifications. A recommended, improved classification system recognizes that the sorbent and SO_2 product can vary independently and it provides for two-way variation. Processes are classified by type of SO_2 product (throwaway or recoverable in the U.S.) and by type of sorbent (nonregenerable or regenerable sorbent for SO_2 and sorbent for an SO_2 derivative such as SO_3). The form of sorbent (solution, slurry, solid, etc.) is applicable to each of the six primary categories.

The development levels of the principal types of FGD processes are shown in Table S.1. Japan has been the most active country in installing commercial FGD (12,000 MW) in power plants. A U.S. process (W-L) had its first power plant application in Japan, and a commercial Japanese process (Chiyoda 101) has been tested in the U.S. In Japan (but not in the U.S.) FGD processes are oriented to oil fuels, open-loop water circuits, and a commercial market for the byproducts. West Germany advanced many processes 5-10 years ago and now has three installations at 40 MW or larger. Its novel Saarberg-Holter process appears attractive at 40-MW scale. Except for a single installation in England (a tall-stack proponent), other countries do not have commercial FGD in operation in power plants. The USSR has a long record of FGD-related development, including fertilizer production from industrial SO_2 tail gases; however, their application to power plants has thus far been limited.

FGD development in the U.S. has been complex in its mix of participants, programs, and funding. U.S. Government participation became strong in the late 1960's and intensified in the early 1970's as air quality control standards gained in legal standing. In addition to the in-house and contracted development by EPA and other agencies, the primary development came from commercial organizations, such as process and equipment vendors,

Table S.1. Development Level of Principal FGD Process Types

Process Type	Vendor ^b	Development Level ^a			
		U.S.	Japan	W. Germany	Other
<u>Throwaway Product</u>					
Nonregenerable Sorbent:					
Solution -					
Sodium Carbonate	CEA	C			
Slurry -					
Limestone	Several	C	C		C
Lime	Several	C	C	C	
Regenerable Sorbent:					
Solution -					
1% Sulfuric Acid	CY 101	PR	C		
Sodium Solution	CEA/ADL, FMC, EN	PR,D*,C*			
Sodium Sulfite	Showa - Denko		C		
Lime - Calcium Chloride	Saarberg - Holter			PR	
Basic Aluminum Sulfate	Dowa		PR		
Sodium Acetate	Kureha		P		
Solid -					
Activated Carbon	Hitachi		C		
<u>Recoverable Product</u>					
Regenerable Sorbent:					
Solution -					
Sodium Citrate	BOM	P,D*			
Sodium Phosphate	Stauffer	P			
Sodium Sulfite	W-L	D,C*	C		
Sodium Carbonate	AI	P,D*			
Ammonia	CAT/IFP	P			PR
Potassium Carbonate	CONSOL	P			
Glyoxylic Acid	Spring - Nobel Hoechst				P
Slurry -					
Magnesium Oxide	UE	D,C*			
Iron Sulfide	PENSYS	P			
Solid -					
Activated Carbon	FW-BF	PR,P	D	PR	
Copper Oxide	Shell/UOP	P	PR		
Melt -					
Sodium-Potassium-					
Lithium Carbonate	AI	PR			

^a C - Commercial
D - Demonstration
P - Pilot Plant
PR - Prototype
* - Under construction or planned

^b Vendor codes are given in Appendix B.

commodity suppliers, utilities, and industrial boiler users. More recently, the Electric Power Research Institute (EPRI) has become increasingly active in FGD development. Through the nature of EPRI's sustaining membership, this participation brings FGD development much closer to its area of application.

FGD development has been conducted at laboratory, bench, pilot-plant, prototype, and demonstration levels. Its record of success has been higher when an appropriate level was used for progressive development rather than for success-failure testing. Successful pilot-plant programs have been run by TVA and by some vendors and utilities; productive prototype operations are illustrated by the Shawnee and by part of the Scholz programs. Most demonstration-scale tests were started prematurely, funded inadequately, and stopped short of complete success. Also, all FGD plant installations prior to 1973 should be considered developmental. Their "operating" experience shows that short-life, part-time operation, and/or a succession of major process and equipment changes reduced their performance below that of a production level to "development-during-operation." The full-scale tests of 1968-72 (demonstration and regular plant) were difficult, expensive, and failure prone and they brought FGD into stronger disfavor.

Commercial U.S. FGD installations are listed by process type in Table S.2. Virtually all currently operational FGD in the U.S. uses limestone or lime slurry processes. The 21 FGD installations (6900 MW) comprise 8 limestone slurry units (3600 MW total), 10 lime slurry units (3000 MW total), and 3 Na_2CO_3 solution units (375 MW total). All of the above installations have a throwaway SO_2 product and a wet, once-through sorbent which includes alkaline flyash in four of the lime-limestone units.

Of the 21 current installations, the first startup of a surviving and reasonably operational system was that of La Cygne in February 1973. Because of the 4-5-yr lead time, the designs for even the most recent FGD startups (mid-1977) were selected prior to 1973 sustained operating experience. This means that the essentials of all FGD systems now in operation were unproven firsts and that commercial FGD is about to complete its first full cycle of development-design-startup-operation-modification.

In terms of sulfur content, operating experience is even more limited as shown in Fig. S.1. Ten of the 21 operational units burn low-sulfur coal

Table S.2. Capacity by Process of Commercial FGD Installations in the U.S.^a

Process Type	Sorbent	Operational		Under Construction Plus Planned ^b		Total	
		Units	MW	Units	MW	Units	MW
<u>Throwaway Product</u>							
Nonregenerable:							
Solution	Na ₂ CO ₃	3	375	2	675	5	1,050
Slurry ^c	Limestone	8	3,580	25	9,700	33	13,280
	Lime	10	2,970	15	8,100	25	11,070
Regenerable:							
Solution	Na ₂ CO ₃	0	0	2	825	2	825
<u>Recoverable Product</u>							
Regenerable:							
Solution	Na ₂ CO ₃	0	0	2	715	2	715
Slurry	MgO	<u>0</u>	<u>0</u>	<u>1</u>	<u>240</u>	<u>1</u>	<u>240</u>
<u>Total</u>		21	6,920	47	20,260	68	27,200

^a Excluding demonstration, prototype, and smaller.^b Contracts awarded plus letters of intent, excluding "requesting/evaluating bids" and "considering FGD system."^c Includes lime-limestone/alkaline flyash systems.

FIG. S.1. SULFUR LEVEL IN COAL AT OPERATIONAL FGD INSTALLATIONS IN U.S. POWER PLANTS
 -(STATION CODING AS IN TABLE 6.2 OF TEXT)

but 5. units are still undergoing extensive modifications or have less than 1 yr of operation. The five medium-sulfur coal users are also in a state of change or have less than 1 yr of operational experience. Of the six high-sulfur coal users, only three have more than 1 yr of experience; namely, La Cygne since February 1973, Green River (64 MW) since September 1975, and Bruce Mansfield (high to medium sulfur) since April 1976. Since the problems associated with FGD escalate with the sulfur content of coal, FGD experience at plant scale needs to be considered in the context of the above profile.

The 28 commercial FGD installations now under construction in the U.S. have scheduled startup dates from mid-1977 into 1980; the 19 formally planned units have less definite startup dates extending into 1982 and 1983. Together, these units comprise 25 limestone slurry units (9700 MW total), 15 lime slurry units (8100 MW total), and 7 units (2500 MW total) distributed among Na_2CO_3 solution (throwaway), two-stage Na-lime (throwaway), and two processes with recoverable SO_2 product (W-L/AC and MgO). As can be seen, lime-limestone processes continue to predominate.

Over the past 10 yr FGD process trends at plant scale in the U.S. can be followed by an evaluation based on a combination of product and sorbent classification. Within the category of throwaway SO_2 product and non-regenerable sorbent, sorbents have moved from solid (obsolete limestone addition to boiler) to solid/slurry scrubbing (obsolete lime addition with scrubber), to slurry (sorbent added to scrubber loop in several limestone and lime slurry processes), to solution (Na_2CO_3). The next movement by category was to throwaway SO_2 product and regenerable sorbent (two-stage or double-alkali processes with one of several soluble sorbents and limestone or lime to precipitate the SO_2 product as sludge or crude gypsum and to regenerate the sorbent). The latest major movement by category has been to recoverable SO_2 product and regenerable sorbent. The commercial examples in that category are W-L/AC with a soluble sorbent and MgO with a slurry sorbent. Other process trends are at various stages of noncommercial development.

Operational FGD plant experience continues to reflect a number of problems and overall progress has been slow, difficult, and expensive. Especially with high-sulfur coal, limestone and lime FGD systems have experienced severe buildups and plugging in the scrubbers and mist eliminators, as well as widespread erosion and corrosion of equipment. Frequently the quality of

equipment supplied has been disappointing. Operating availability has been maintained only by cleaning, washing, and heavy maintenance. (In general, the construction materials have been less than satisfactory for service with slurries.) Using a variety of measures to keep systems running, availabilities have gradually risen above 90% at some installations, but have remained unsatisfactory at others.

Improvements with limestone and lime systems have been made in design, equipment, and operating procedures. In most current installations, partial or complete particulate removal has been combined with SO₂ removal, and in general, the degree of both particulate and SO₂ removal has been adequate for emission standards. Most new plants will use conforming ESPs for particulate removal. More resistant alloys and coatings are being tried and wear plates are being used at critical locations. ID fans are cleaned often to avoid vibration problems. For both slurry and solution service, there is a growing trend toward spray scrubbers and more effective types and locations of their sprays are reducing buildups. Mist eliminators have varied from effective to fragile and nonfunctional. Their design, performance, and effective washing are often unreliable. Flue gas reheat, a consequence of gas cooling by wet scrubbing, is always costly and several plants have experienced nagging corrosion with the steam-heated, inline, heat exchanger type. At the same time, other inline reheaters have held up well and the reasons for this disparity are not clear. Sludge disposal from limestone and lime systems is a continuing problem. Requirements for pond integrity are being tightened, sludge fixation processes are being installed at some plants, and sludge oxidation to a more settleable sulfate form is now being tested and pursued with vigor.

FGD costs remain confusing, debatable, and highly site-specific. When particulate and SO₂ removal have been combined, there has been a tendency to associate the total cost with SO₂ removal. The current price of FGD systems is not always compared with the current price of power plant construction. Importantly, the full scope of limestone and lime systems has finally become apparent in comparison to its so called "simplicity" of 5-10 yr ago. Such variables as redundancy, retrofitting, and cost escalation can increase the capital investment for a limestone or lime system from as little as \$75/kW to as high as \$125/kW. Annual revenue requirements (which include cost of

capital) can range from about 3-5 mills/kWh. Inclusion of sludge fixation raises the latter by 1/2 to 1-1/2 mills/kWh and makes limestone and lime processes more costly than some recovery processes receiving byproduct credit.

Beyond the W-L and MgO systems, which are scheduled for commercial operation, there is a conspicuous lack of development with recovery-regenerable processes. Most of the usually discussed 8-10 processes exist at pilot-plant or smaller scale and they do not have a data base for confident evaluation beyond the next development level. Evaluations projected to 100 MW can, therefore, be deceptive. From many standpoints -- costs, resources, land conservation, etc. -- these processes offer much improved long-term prospects for FGD but their advance to commercial status will require major development and assistance comparable to that given to the throwaway types.

Limestone and lime have achieved their current position based on 10 yr of effort. This momentum is continuing and future improvement can be expected to help meet rising requirements. However, some problems with limestone and lime have had sufficient impact so that the two-stage or double-alkali type is appearing as a potential alternative. The performance of its scheduled commercial units should have a strong influence on the course of FGD.

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OF GENERATING ELECTRIC POWER FROM COAL

1977 TECHNOLOGY STATUS REPORT

*Appendix F: Flue Gas Desulfurization
in the United States - 1977*

Prepared by
Tennessee Valley Authority

ABSTRACT

This report details the current status of flue gas desulfurization (FGD) in conventional coal-fired power plants in the U.S. Specifications and features are listed for all commercial FGD installations now in operation or being constructed in this country. The FGD operational experience in each plant is summarized and evaluations are made of process, equipment, and operational performance. FGD in other countries, mainly Japan and West Germany, is reviewed for applicability in the U.S.

Federal and state air quality control standards are described and their prime influence on FGD development and plant installation is noted. The rising importance of coal is discussed and some limitations on low-sulfur coal are included. Other potential approaches to desulfurization, such as fluidized-bed combustion, are discussed as far as relevant to FGD.

Limestone and lime slurry processes greatly predominate among existing and planned FGD installations in the U.S. However, soluble SO₂ sorbents are noted as entering use for both throwaway and recoverable SO₂ products. It is recognized that recovery processes offer attractive long-term prospects in terms of resources and of apparent cost of operation. However, they remain very limited, both in number of processes and in level of development.

The report shows that sustained commercial FGD operation in the U.S. is very young, and that process and equipment experience only now is starting its second cycle. Additional progress can be expected as knowledge is derived from sound development programs and from operating experience at recent installations.

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1 INTRODUCTION

This study deals with the current status of flue gas desulfurization (FGD) as a control technology for conventional coal-fired power plants. It is concerned primarily with FGD technology used in the U.S. but includes U.S. processes which are being applied abroad and some outside processes of concern to this country. In all processes FGD deals with the capture and removal of sulfur dioxide (SO_2) after combustion.

Over the years the status of FGD has been controversial and its alternatives have been discussed as if they were available without limitation. It will be seen that alternatives, such as low-sulfur fuel and coal cleaning, are being utilized for many power plants, but that other more sophisticated techniques are for the future. The long-term potential of more exotic alternatives often is compared with the current capability of FGD but the time required for their development and construction will be as real as it has been for FGD. In addition, the current capabilities of FGD can be improved. Fluidized-bed combustion and other potential approaches to environmentally acceptable electrical generation do not fall within the current subject but their development positions are of obvious interest as sulfur control alternatives.

Throughout the past decade, claims and counterclaims have questioned whether FGD had reached a maturity of commercial acceptability. It is clear that protagonists have based their claims on widely differing criteria or standards of acceptability. What satisfied one participant could be grossly inadequate for the needs of another. In view of the complex situation, this review judges FGD acceptability based on fully operational plants and categorizes demonstration units as developmental. Some history and background are included but their comprehensive treatment would be beyond the present scope -- the references provide some of these extensions.

Advances in FGD have come from and by many sources, but during the decade of about 1965-75, the course of FGD development came under strong influence by the type and amount of U.S. Government-sponsored research and development. The traditional methodology of industrial development was changed from a profit orientation to one of necessity. During the same period, pollution standards were being set and applied by federal, state,

and local authorities and power plants were confronted with a new type of activity which included heavy chemicals and other processing. An unfortunate schism developed between the regulatory areas of government and this industrial segment, and to some extent is a reason for the claims and counter-claims as to the acceptability of FGD as a means of control.

As FGD systems were built, the electric utility industry underwent change at all levels. Power plants enlarged operating and maintenance activities. They increased environmental staff, work, and functions which often modified the production priorities in the plant. The local personnel were obligated to making the FGD installation work. Corporate activities were enlarged or added in environmental, legal, and financial areas and in managing the changing economics of a utility company. The nature of the utility business was changed. Building and operating permits became complicated in both content and attainment. Approval of rates often was made conditional upon compliance with pollution abatement, and accounting conventions were adjusted to meet the new profile of economics.

It is important to realize that FGD has developed in troubled and often confused circumstances. Now, however, FGD may be at a stage which permits better assessment and less contention, and regardless of direction, its future may be based on firmer ground. This improved and clearer status can help to bridge the problems between environmental controls and the increased use of coal in power plants.

2 DESULFURIZATION NEEDS IN THE U.S.

2.1 AIR QUALITY CONTROL STANDARDS

2.1.1 Background¹⁻³

The legal regulation of air quality is the specific reason for the adoption of FGD by power plants. Now, each coal-fired power plant, existing and new, must find a combination of fuel and/or desulfurization which produces a legally acceptable SO₂ emission from the plant. Since the emission standards define one boundary of desulfurization, they are involved directly in its evaluation. The present study is concerned with these standards from the standpoint of what emission levels an FGD system can or needs to meet, but it does not recommend what the standards should be.

The strongest surge in the control of power plant stack emissions came in 1971 when the U.S. Environmental Protection Agency (EPA) set both ambient standards for atmospheric pollutants, and emission standards (standards of performance) for new power plants and other industrial installations. The setting of these standards was based on the Clean Air Act as amended (1970), but this act had been preceded by the Air Quality Act of 1967 and the Air Pollution Control Act of 1955. The intervening 15-yr period had seen accelerated federally sponsored programs in research, training, and process development in the fields of air quality and pollutant capture. Other organizations also had worked for many years, and often heavily, on these subjects. The accumulated body of technology gave some background and became the basis for the 1971 standards. However, capability of applied control equipment was not established at that time.

That year, EPA announced National Ambient Air Quality Standards (NAAQSs) for six pollutants -- particulate matter, SO₂, nitrogen dioxide (NO₂), carbon monoxide (CO), hydrocarbons, and photochemical oxidants. For each pollutant, two standards were issued for ambient conditions in the atmosphere. A primary standard was set at a level considered necessary to be noninjurious to public health; a secondary standard was designed to protect against crop damage, corrosion of materials, reduced atmospheric visibility, and the like.

Two measures were taken to achieve the designated ambient levels. On the one hand, EPA issued Federal Standards of Performance for New Stationary

Sources (often called New Source Performance Standards or NSPSs) which define the maximum emission levels for new power plants in the U.S. Second, EPA obtained from each state a State Implementation Plan (SIP) for local achievement of the EPA standards. When approved by EPA (mostly in 1972), the SIP prescribed local emission levels, timetables, etc., for achieving the ambient levels of NAAQS. For large, new power plants, the SIP is no less stringent than the EPA emission standards, but for existing plants, they vary widely in type and level. Some local agencies impose additional requirements which are not included in this report.

2.1.2 EPA Standards of Performance

NSPSs apply to "new" fossil fuel-fired steam generators with more than 250 MBtu/hr (M = 1 million) heat input to the boiler (about 25-30 MW). For this purpose, "new" plants are those whose construction or pollution-increasing modifications started after August 17, 1971. The standards are shown in Table 2.1. The standard for particulates is of direct concern to FGD when plants have overlapping systems for the removal of particulates and SO₂.

Published literature often is based on the NSPS emission level of 1.2 lb SO₂/MBtu heat input to the boiler. The use of this single value can give the impression that it is a representative or even a sufficient criterion but in fact, the SIP for new or existing power plants can depart far from it.

2.1.3 SIP Regulations

Each state has prepared a SIP which includes standards and timetables for particulate and SO₂ control. The plans are now EPA approved and enforceable, except in Ohio where they are being resolved currently. However, they are not static. As conditions change and as more information is obtained, some states propose revisions which create a shifting situation between current and pending requirements.

In any state the basis for expressing the allowable SO₂ emission rate involves one or more of nine units (e.g., lb SO₂/MBtu for each fuel); applied to one of three settings (e.g., an individual boiler), with a stated time of averaging (e.g., 2 hr). Within a state the emission rate can vary according to location, type of fuel, and size of power plant (e.g., Appendix A for New Jersey). Across the country there are several hundred levels of control and

Table 2.1. EPA Standards of Performance for New Fossil-Fuel Fired Steam Generator⁴

Source Category	Affected Facility	Pollutant	Emission Level	Monitoring Requirement
Subpart E:				
Steam generators (>250 MBtu/hr)	Coal-fired boilers	Particulate	0.10 lb/MBtu	No requirement
		Opacity	20%	Continuous
		SO ₂	1.2 lb/MBtu	Continuous
		NO _x	0.70 lb/MBtu	Continuous
		(except lignite and coal refuse)		
	Oil-fired boilers	Particulate	0.10 lb/MBtu	No requirement
		Opacity	20%; 40% 2 min/hr	Continuous
		SO ₂	0.80 lb/MBtu	Continuous
		NO _x	0.30 lb/MBtu	Continuous
	Gas-fired boilers	Particulate	0.10 lb/MBtu	No requirement
		Opacity	20%	No requirement
		NO _x	0.20 lb/MBtu	Continuous

the practical approach simply is to look up the control standard for the specific power plant or locality -- EPA publications provide such listings.^{2,3,5} The most up-to-date information is obtained directly from EPA or the appropriate state regulatory agency. A similar pattern and procedure exist for particulate emission standards.

The experience of the Tennessee Valley Authority (TVA) shows that a power plant does not have the option under current law to meet the EPA ambient standard, at the neglect of the SIP standard for that location (in this case, an emission rate). TVA had developed a program for meeting the ambient standard by a combination of tall-stack dispersion and adjustment of plant operating level according to atmospheric conditions. However, a U.S. Supreme Court decision in 1976 confirmed the primacy of the emission regulation.⁶

2.2 FUELS FOR ELECTRICAL GENERATION

2.2.1 Fuel Needs for Electrical Generation

Fig. 2.1 gives a representative, 10-yr forecast of the fuel for electricity, using data from the National Electric Reliability Council. Total electric generating capability is shown as growing at 5.3%/yr (68% for the decade of 1975-85), coal use at 6%/yr (75% for the decade), and nuclear power at 18%/yr (530% for the decade). Natural gas use is seen as peaking in 1977-78 and oil as almost leveling by 1985. Oil and natural gas together would be level in 1981-83 and trend downward in 1984 and 1985. The "other" category in Fig. 2.1 is mainly pumped storage but it includes the very small amount of geothermal power. Exotic fuel such as solar energy captured in solar cells (distinct from solar heat) is commercially far in the future.

Fig. 2.2, based on the same data, shows the percentage of total generating capacity provided by each fuel. As in Fig. 2.1, coal's share would equal that of oil plus gas in 1977. During the decade, coal's share would increase only from 38.5 to 40.2% in spite of its 75% increase in electrical equivalent (not necessarily in coal tonnage) in Fig. 2.1. The shares of oil and of natural gas decline steadily in the forecast but nuclear power's fraction increases from 7 to 22%. This means that, omitting the small other category, nuclear and coal are the only fuels with growing shares in Fig. 2.2. Thus, coal and nuclear power emerge as the main domestic prospects for the projected levels of electrical generation.

FIG. 2.1. ESTIMATED ELECTRICAL GENERATING CAPABILITY IN THE U.S. (DATA FROM REF. 7)

FIG. 2.2. PERCENTAGE DISTRIBUTION OF FUELS
IN ESTIMATED ELECTRICAL GENERATION (DATA FROM REF. 7)

Several factors are firming or enlarging the position of coal, and potentially of FGD, in the mix of fuels forecast for electrical generation. In late 1976 the Federal Energy Administration (FEA) and EPA began directing that 74 units at 32 existing power stations be scheduled for changeover from oil or gas to coal.⁸ For plants now designed or under construction, the latest oil- or gas-based plant startup is in 1979 and the latest oil-fueled plant startup is in 1982.⁹ Future fossil-fueled power plants are required to be designed with coal-burning capability. Also of future effect, the new orders for both nuclear and fossil-fueled power plants virtually collapsed in 1975 and 1976.^{7,9,10} These falloffs will lead to less gain in electrical capacity in the mid-1980's or to an additional increase in coal-fired plants which require less construction lead time than nuclear.¹⁰

2.2.2 Fossil Fuels for Electrical and Total Energy

The types and amounts of fossil fuels used for electrical generation (and hence of FGD concern) have been determined almost directly by their supply-demand-price situations in the country's total primary energy picture. Fig. 2.3 shows how electrical energy generation fits into the energy flow pattern for the U.S. On a heat-equivalent basis in 1970, electrical generation used 50% of total coal consumption, 18% of total gas, and 7% of total oil. At that time, fuels for electricity accounted for about 22% of the fuels for total primary energy consumption -- by 2000 they could reach 50%.¹¹

Primary energy growth since World War I has come from oil and natural gas (Fig. 2.4). Starting about 1918, crude petroleum production began a rapid expansion whose peak in 1970 is not expected to be equalled in 1977 even including oil from Alaska's North Slope.¹³ Natural gas production had a similar but greater growth, beginning strongly in the 1940's. It peaked in 1973.¹³

On the other hand, coal has had no similar growth since about 1918. It has achieved a very modest long-term growth in tonnage but the heating value of bituminous coal and lignite production has remained essentially constant in the band of about 10,000-15,000 TBtu/yr (T = 1 trillion) (Fig. 2.4). Anthracite coal, small in proportion, declined during that period. Since total energy production (or consumption until the 1950's) continued to

Source: "Understanding the National Energy Dilemma," JCAE, 1973

FIG. 2.3. U.S. ENERGY FLOW PATTERN - 1970

FIG. 2.4. PRODUCTION AND CONSUMPTION OF ENERGY IN THE U.S. ¹²

increase at about 3%/yr, the percentage share of bituminous coal and lignite fell from the 70's to about 22 (Fig. 2.5). The percentage share of total coal, including anthracite, fell from about 90 to 22 during the present century. It should be emphasized that coal's lack of production growth since World War I has stemmed from lack of growth among uses rather than from inability to mine more coal. In one application after another, users preferred other fuels.

The U.S. Bureau of Mines (BOM) statistics for 1976 show that electric utility use of total fossil fuels increased 7.2% over 1975.¹⁴ Coal provided 59% of the fossil fuel input. Oil use in electrical generation rose by 7.5% but natural gas use declined by 2%. Consumption of total primary energy has risen again, virtually to its 1973 level, but the energy to gross national product ratio is continuing its downward trend begun in 1971. Net imports of crude petroleum products reached 40.6% of petroleum demand for 1976. This means that net imports amounted to 68% of net domestic production.

Fig. 2.4 shows that the deficit in total energy became severe after 1970 when production increases almost stopped. This leveling of production followed the earlier (1964-69) curtailments of interstate natural gas prices and reduction of tax allowance for depletion of oil and gas reserves. On the other hand, oil consumption began to rise sharply after 1970 when the environmental upsurge was demanding lower emissions from power plants. Since FGD capability had not been proven at plant scale, the emissions "solution" was obtained by widespread switching from coal to low-sulfur oil. The oil deficit, created by the leveling production and compounded by the increasing demand, was made up mainly by successively larger oil imports. This set up the vulnerability to the oil embargo of 1973 and to the price escalations which have grown out of bounds. Domestic coal prices have increased in phase, though not in the same proportion as oil.¹⁵ To the extent that coal is made the relief to the above dilemma, FGD or its sulfur-control alternatives must meet the backlog as well as the currently nonconforming and future needs.

2.2.3 Coals for Electrical Generation

Coal deposits occur widely in the U.S. (Figs. 2.6 and 2.7), but 64% of cumulative production through 1973 has come from the bituminous beds of Pennsylvania, West Virginia, and Kentucky (Fig. 2.8). Adding Illinois

FIG. 2.5. PERCENTAGE SHARE OF PRIMARY FUELS IN U.S. ENERGY PRODUCTION ¹²

FIG. 2.6. COAL FIELDS OF THE CONTERMINOUS U.S. ¹²

FIG. 2.7. REMAINING IDENTIFIED COAL RESOURCES OF THE U.S.-JANUARY 1, 1974 ¹²

FIG. 2.8. PERCENTAGE DISTRIBUTION, BY STATES,
OF CUMULATIVE COAL PRODUCTION IN THE
U.S. TO JANUARY 1, 1974 ¹²

bituminous coal raises the total to 74%. Most of this coal has been used by utilities located at or near the coal occurrence. By 1973 they accounted for 70% of total bituminous coal and lignite consumption in the country (Table 2.2).

In recent years however, two factors have sharply increased interest in western coal, especially in the subbituminous beds of Montana and Wyoming. The power plant capacity in the Far West has increased, but more importantly, the low-sulfur content of subbituminous coal has appeared to be a way of avoiding desulfurization -- especially FGD. To meet this prospective demand, Montana and Wyoming mines are being developed so massively that some are falling far behind schedule.¹⁶ Even so, total western coal in FY 1975 supplied 55% of the U.S. electric utilities' coal of less than 1% sulfur; it provided 17% of their total supplies (Fig. 2.9).

Very large reserves of lignite occur in North Dakota and Montana, with small reserves in Texas (Fig. 2.7). Although low in sulfur, it is also low in heating value. In equivalent heat and hence as a limiting transportation factor, 2.0 tons of lignite or 1.4 tons of subbituminous coal correspond to 1 ton of bituminous coal.

Table 2.2. Percentage Distribution of Bituminous Coal and Lignite in the U.S.¹²

Consumer Class	1943	1953	1963	1973
Utilities	12	26	51	70
Steel industry				
Coke production	17	26	19	17
Steel and rolling mills	2	2	2	1
Manufacturing industries	26	25	22	11
Retail deliveries	21	14	6	1
Railroads	<u>22</u>	<u>7</u>	<u>Neg</u>	<u>Neg</u>
Total	100	100	100	100

W : WEST (ROCKY MOUNTAIN AND NORTHERN GREAT PLAINS STATES)
 C : CENTRAL (INCLUDING WESTERN KENTUCKY AND TEXAS)
 E : EAST (OHIO, EASTERN KENTUCKY, AND APPALACHIA)

FIG. 2.9. ORIGINS OF FISCAL YEAR 1975 COAL SUPPLIES TO ELECTRIC UTILITIES BY RANGES OF SULFUR CONTENT 17

2.3 SULFUR CONTROL ALTERNATIVES

2.3.1 Low-Sulfur Fuels

In concept, the simplest option for sulfur control is a fuel whose sulfur emission does not exceed the air control standard. Since natural gas is virtually sulfur free, its use has avoided sulfur control problems as well as simplifying fuel handling and combustion. Because of supplies, it is now a limited option. Fuel oil within the conforming sulfur level also has been used widely and without FGD. Nonconforming oil involves FGD which is conventional in sulfur removal but very low in particulates involvement. When fuel supplies vary around the conforming sulfur level, blending of higher and lower sulfur fuels can increase the tonnage of conforming fuel. By this method, oil delivered to steam-electric plants in 1973 was raised from 70 to 73% conforming; coal was raised from 42 to 48% conforming.¹⁸

Low-sulfur coal has substantial but not universal potential as a sulfur-control alternative. Table 2.3 shows that each type of coal has some deposits at low sulfur and that the subbituminous rank has nominally 99.6% of its deposits within 1% sulfur. However, many combinations of SIPs and coal heating values can lower the compliance level of coal to much less than 1% sulfur.^{19,20} Fig. 2.10 shows the approximate relationship among these items. (It neglects the small proportion of sulfur going to boiler bottom ash and to some other exits.) A subbituminous coal of 9500 Btu/lb and 1% sulfur would satisfy a SIP of 2.1 lb SO₂/MBtu. A SIP of 1.2 would require a coal sulfur

Table 2.3. Percentage Distribution of Identified U.S.
Coal Reserves According to Sulfur Content¹²

Rank	Sulfur Content (%)		
	Low	Medium	High
	0-1	1.1-3.0	3+
Anthracite	97.1	2.9	-
Bituminous coal	29.8	26.8	43.4
Subbituminous coal	99.6	0.4	-
Lignite	90.7	9.3	-
All ranks	65.0	15.0	20.0

FIG. 2.10. NOMOGRAPH RELATING SULFUR CONTENT AND CALORIFIC VALUE TO EMISSION STANDARD ²⁰

of 0.57% and a SIP of 0.6 would require a coal sulfur of 0.28%. The break-point for "low-sulfur" coal therefore is often in the tenths of 1% rather than at 1%. On the average, the 1.2 standard could be met, without controls, by about 54% of the Montana deposits and by about 57% of the Wyoming deposits.²¹

The use of low-sulfur western coal (or of any basic change in coal) involves three areas of cost consideration, namely cost of delivered coal, cost of power plant conversion, and indirect costs coming from power plant derating and from operating problems.

As with other commodities, the cost of western coal at the mine will find its competitive level. For out-of-state shipment, the Wyoming tax on coal has tripled in 2 yr to 10.4% of the selling price, and the Montana tax is 30%.²² Transportation cost soon exceeds the f.o.b. mine cost. When the f.o.b. mine cost of low-sulfur Wyoming coal was \$7/ton, the delivered cost per heat-equivalent ton was \$36 in East St. Louis and \$45 in Philadelphia.²³ However, partial offsets for the utility occur in the avoided desulfurization costs and in the pass-on nature of coal clauses in the electricity rate schedules.

Additional equipment costs for burning low-sulfur coal come mainly from differences in coal handling, in the combustion system, and in particulate collection. Though highly variable, these capital costs for conversion have been approximated at \$20/kW.²³ Normally, they are higher for modified than for originally installed systems.

Among the equipment modifications, electrostatic precipitators (ESPs) for particulate collection are of special importance. Their design is highly specific to the application²⁴ and in power plants it has depended on the built-in conditioning of flyash particles with small but definite amounts of sulfur trioxide (SO_3). (In other respects and in some FGD situations, SO_3 is undesirable.) Figs. 2.11 and 2.12 show the subtle reasons for this effect. Fig. 2.11 shows that the precipitation rate in an ESP drops abruptly with increased resistivity of the flyash particles. In Fig. 2.12 low-sulfur coal is one cause of high resistivity. The low-sulfur coal produces less SO_3 and hence less conditioning of flyash particles toward lower resistivity. The practical effect of this phenomenon in a power plant which had changed to low-sulfur coal in 1971 was a drop in collection efficiency from 95% to below 50%.²⁶

FIG. 2.II. DEPENDENCE OF ELECTROSTATIC PRECIPITATION RATE ON PARTICLE RESISTIVITY ²⁵

FIG. 2.12. DEPENDENCE OF FLYASH RESISTIVITY ON SULFUR IN COAL AND ON FLUE GAS TEMPERATURE ²⁴

A similar effect was one reason for the failure of an FGD approach when the furnace-injected limestone contained magnesium (MgO) compounds which removed SO_3 by reacting with it.²⁵

The above problem in ESPs is largely correctable by increasing precipitator size, by adding conditioning agents to the flue gas, or by using high-acceleration rapping of the precipitator plates.²⁷ However, its cost of correction is one of the costs of converting to low-sulfur coal.

The third category, indirect costs from plant derating, etc., can be seen in the description of a 1971 change from Illinois to western coal.²⁶ Due to high moisture, an induced-draft (ID) fan's capacity was reduced by 20% and total plant derating was greater than 20%. A derating of 20% increases costs/kW by 25%. This description of changing coals²⁶ includes the efforts needed in the several segments of coal shipment, handling, and combustion, as well as those for flue gas treatment. Even though the conversion of an existing plant can be difficult, expensive, and time consuming, these difficulties do not preclude the use of western coal.

2.3.2 Fuel Desulfurization

Beneficiation of coal by physical means covers a wide range of process treatment and of type of results. One classification of physical coal cleaning²⁸ describes six degrees or intensities which range from no treatment to deep cleaning for pyrite (FeS_2) removal. Coal crushing, washing, and drying are applied increasingly in the second through fifth levels, mainly to remove clay and ash. Some pyritic sulfur may be removable at the fourth and fifth levels of treatment. A degree of physical coal cleaning at these levels has been used for decades, and it is generally available in terms of the technology and the equipment.

The sixth level of physical coal cleaning²⁸ adds deep cleaning of at least part of the coal by heavy-media and other separations. It differs from the previously mentioned methods in its complexity, its status of development, and by its ability to remove up to about 85% of the pyritic fraction of the coal's sulfur. (Depending on the particular deposit and seam, the pyritic sulfur may comprise from a few to roughly 50% of the coal's total sulfur.) In this category, the new General Public Utilities process²⁹ produces a

conforming low-sulfur coal fraction, a higher sulfur fraction to be used with FGD or in plants with higher SIP regulations, and a refuse fraction which contains the eliminated FeS_2 . This process is being tried at the demonstration stage to improve the combination of sulfur removal, operating costs, and coal losses. For the country as a whole, some minority of the coal reserves (13-32%) may be amenable to meeting the NSPS requirement through physical cleaning.^{11,12,20}

Chemical coal cleaning removes organic as well as inorganic sulfur compounds. Many process developments are underway using solvents (inorganic or organic), hydrogenation, and various combinations of temperature and pressure. The resulting fuel can differ considerably from coal. The stages of development range from laboratory to prototype in size and the results from demonstration-scale testing are not likely to be stabilized before the early 1980's.

Coal gasification and liquefaction produce fuels quite different from coal and often these processes are termed coal conversion. They include desulfurization but they represent much more than alternatives for sulfur control. Their readiness for commercial use in power plants obviously is some years in the future.

Coal combustion in a fluidized bed, in place of a conventional furnace and boiler, is receiving great development effort which will reach large-scale testing during the early 1980's. Limestone (or lime as CaO) is mixed with the coal or added to the bed to capture SO_2 during or at the point of combustion. The SO_2 -lime product joins the ash for removal. From the desulfurization standpoint, the limestone (or lime) sorbent for SO_2 travels once through and its sulfur product is for throwaway, although lime regeneration possibilities have been proposed and are being developed. Regardless of desulfurization, the process has inherent features with attractive potential. Reaction rates are relatively fast, and the bed temperature is significantly lower than in conventional boilers. This could lower the production of nitrogen oxides (NO_x) which is now being promoted for more stringent control. Heat transfer rates are unusually high in fluidized beds. Importantly, a background of partially applicable experience began about 1950 in the fluidized combustion of metal sulfides in the same range of high temperature and with difficult bed materials. The fluidized bed now could be viewed as a highly prospective sulfur control alternative to FGD.

2.3.3 *Comparison of Sulfur Control Alternatives*

A basic feature of the status of FGD is its place among sulfur control alternatives. Low-sulfur coal and coal desulfurization are the main alternatives to FGD which are available for current use. Each has its limitations and the deep-cleaning types of physical coal cleaning now are receiving additional development. The other methods of coal desulfurization are at various stages of development. Low-sulfur coal was shown to be applicable, by itself, to many but not all power plants.

It will be developed later that FGD has progressed past its struggle of the early 1970's and now has a position which is operational but still needing much improvement. Often, FGD is used with low-sulfur coal and there are some situations where it is likely to be used with coal that has had considerable physical cleaning.^{23,30} Since each method has its areas of capability and of limitation, the combination of costs and environmental effects will determine their mix.

3 WORLD FGD PROCESSES

3.1 HISTORY

3.1.1 *Early FGD in England*³¹⁻³³

After more than 40 yr, the FGD developments at the Battersea and Fulham power plants in London remain important in the history of FGD. The processes, the perception of process parameters, the development of operating techniques, the successes, and the problems all stand as forerunners to the current decade of FGD in the U.S.

FGD at Battersea resulted from severe atmospheric stagnation in 1929. The process was developed by the London Power Company for dust and SO₂ control at its Battersea and later at its Bankside station in London. Including revisions, this process of the early 1930's had cyclone collectors for particulates, once-through washings with alkaline water from the Thames River, partial pH control by the addition of a chalk /calcium carbonate (CaCO₃) slurry, oxidation by aerators, spray eliminators for the final flue gas, and a neutralizing dilution of the effluent. The process achieved 90-95% SO₂ removal. (In the 1930's the fuel was coal with 0.9% sulfur; recently it has been oil.) The problems, in addition to operations and maintenance, have been the large amounts of water to be handled, the lowering of the river's pH, and an intermittently drooping stack plume from the unheated flue gas. In fact, the two Battersea scrubbers were closed down permanently in 1969 and 1974 due to plume problems. Bankside (300 MW) now has England's only FGD installation.

The Thames River's limited capacity for Battersea-type effluents led to the development of the Howden-ICI (James Howden and Company - Imperial Chemical Industries) lime scrubbing process. Under closed-loop operation, it produced a thickened sludge for barge movement to remote disposal. The process was tried for a time at the Tir John station in Swansea, but soon it was abandoned, overburdened by the high ash in the coal. In 1936 it was installed at the 40-MW Fulham station where it operated until World War II. With England's postwar leaning to tall-stack dispersion, the plant was not rehabilitated.

In the pilot-plant development program which used ash-free flue gas, the scaling potentials of calcium sulfite (CaSO_3) and calcium sulfate (CaSO_4) were found to be controllable by (1) reducing their supersaturation in the recirculated liquor by providing crystallization time (3.5 min) in a holding tank, (2) providing an advantageous surface for residual crystallization in the scrubber in the form of at least 3% of each of CaSO_3 and CaSO_4 crystals in the liquor, (3) using "nonscaling" packing (suspended wooden plates) in the lower part of the scrubber, and (4) using high liquor rates. Forty years later the ICI set of operating parameters (including the above measures) still are useful guidelines for controlling scaling in limestone and lime slurry processes.³²

3.1.2 Highlights of FGD History

In FGD progress the respective decades of the 1950's, 1960's, and 1970's could be labeled as beginnings, development, and plant operation. However, these terms do not imply a coordinated or continual expansion for this long period in the U.S. or in other countries.

Prior to about 1950 FGD development around the world was spotty. In England it reached the previously described plant levels in the early thirties but these examples did not lead to sustained FGD development. In the U.S. H. F. Johnstone (University of Illinois) started a career of research on SO_2 removal around 1930 but much of his work was too early for wide interest. The USSR began a major research effort in SO_2 recovery in the 1940's.³⁴ Also, FGD literature often includes the Cominco Ltd. process, which started in Canada in 1932, for scrubbing nonpower plant concentrations of SO_2 with ammonia (NH_3). Over the years the production and removal of SO_2 and SO_3 had occurred mainly in nonferrous smelters and in byproduct and sulfur-burning sulfuric acid (H_2SO_4) plants. They have a history of improvements in the handling of sulfur gases. Of interest to FGD, improved SO_3 absorption in H_2SO_4 plants eliminated the tail-gas lime boxes around 1930. Under changed economics, FGD seeks about the same proportionate removal of sulfur from flue gases as did H_2SO_4 plants from their feed gases.

The 1950's brought several FGD-related developments. In 1953 TVA began pilot-plant testing of several FGD processes^{34,35} but the program

remained at pilot level because of the apparent adequacy of tall-stack dispersion plus atmospheric monitoring.^{36,37} During the same period (1952), the USSR equipped a 65-MW power boiler for SO₂ removal by NH₃ scrubbing³⁴ and B. A. Chertkov at Soviet Union's NIIOGAZ (Scientific Research Institute for Industrial and Sanitary Purification of Gases) soon started his major contributions to FGD research and technical literature. In Japan, England, and West Germany, a number of SO₂ scrubbing units were installed on H₂SO₄ and other industrial plants.³⁸

The 1960's was the decade of great change and growth in FGD development. Several dozen processes of various types were announced in the U.S., Japan, West Germany, France, and in the Eastern European countries of USSR, Romania, and Czechoslovakia. They ranged from simple proposals or paper processes to those supported by substantial development.

Boiler addition of dry limestone was a major development of that period. It had participation (in addition to power plant participants) by 15 organizations in the U.S., 4 in West Germany, and 2 in Japan.³⁹ As a result of plant trials, the early concept of collecting the SO₂ product in the regular ESP was revised in favor of scrubbers. This combination survived pilot-plant testing at Wisconsin Electric Power Company and at Detroit Edison Company plants.³¹ The plant installations at Union Electric Company, The Kansas Power and Light Company, and Kansas City Power and Light Company (5 units with 12 scrubbers in all) were influential in the standards setting process but were not satisfactory in operation and boiler addition of limestone gradually has been abandoned. In the meantime, obviously premature claims of success had reduced the potential of FGD in many minds. Actually, the failures and plant revisions evolved into limestone and lime slurry scrubbing which now are the major commercial processes in this country.

NH₃, MgO, sodium (Na), and other processes had considerable development activity in several countries during the 1960's⁴⁰ but the major change in the U.S. was the entry of federal agencies into the picture. The principal agencies were the National Center for Air Pollution Control and its successor, the National Air Pollution Control Administration (both in the U.S. Department of Health, Education, and Welfare); EPA succeeded in 1970. The scope, pace, and orientation of FGD development were changed by their in-house work and by

their selective funding (with proportionate management) of contracted work in all phases, from evaluations to laboratory, pilot-plant, and other testing.⁴¹

The 1970's has seen full-scale plant installations as well as development of FGD. Many power plants had been evaluating process offerings but the schedules of the previously discussed air quality control standards brought them to prompt commitment for installations. As discussed later, almost all power plants found much development still ahead of them, even after installation.

On the development side, EPA conducted or funded programs for the evaluation and cost comparison of processes. Some advanced noncommercial processes have been tested at prototype- or demonstration-scale and the lime-limestone types now have programs for process and equipment improvement.^{42,43} These activities are continuing.

In addition to power plants, 119 organizations in 16 countries are active in FGD research, and of these, 62 are in the U.S., 22 in Japan, and 8 in West Germany.⁴⁴ Only the U.S. and Japan have proceeded to more than nominal amounts of commercial FGD on power plants.

3.2 TYPES OF WORLD PROCESSES

3.2.1 Classification

Fig. 3.1 shows the type of classification frequently used for cataloging FGD processes.^{32, 45-48} In it, the throwaway/recovery characteristic has been applied to the process as a whole, and this procedure has led to additional discrepancies as new processes have appeared. In the present report, processes are grouped and discussed in terms of the classification given in Table 3.1. This classification recognizes that the SO₂ product and the sorbent can vary independently and it provides for this two-way variation. On this basis, processes are classified by type of SO₂ product (throwaway or recoverable in the U.S.) and by type of sorbent (nonregenerable or regenerable for SO₂; sorbent for an SO₂ derivative such as SO₃). This arrangement gives six primary categories of FGD processes. In a secondary division, the form of sorbent (solution, slurry, solid, etc.) is applicable to each category and

FIG. 3.1. TECHNOLOGIES FOR THE REMOVAL OF SO₂ FROM STACK GAS ⁴⁵

Table 3.1. Classification of FGD Processes

SO ₂ Product	Sorbent				Form of Sorbent
	A. Nonregenerable Sorbent for SO ₂	B. Regenerable Sorbent for SO ₂	C. Sorbent for SO ₂ Derivative		
I. Thrcwaway	Once-through processes (e.g., limestone) whose SO ₂ sorbents produce disposable products	Processes (e.g., "double alkali") which produce disposable products from SO ₂ sorbed in regenerable sorbents	Processes (e.g., catalytic) which convert SO ₂ , sorb the SO ₂ derivative, and produce disposable products	1. Solution 2. Slurry 3. Solid	
II. Recoverable	Once-through processes whose SO ₂ reactants (e.g., NH ₃) produce commercial products	Processes (e.g., MgO) which produce commercial products using various sorbents, regenerants, and reductants	Processes (e.g., catalytic) which convert SO ₂ , sorb the SO ₂ derivative, and produce commercial products	1. Solution 2. Slurry 3. Solid 4. Melt 5. Gas	

it provides the important solution versus slurry and wet versus dry aspects of sorbents. This new scheme of classification shows consistently where processes are and how they have evolved among groups. It accommodates the important "double-alkali" type, where the SO_2 product is throwaway but the sorbent is regenerable.

Definitions for this classification (Table 3.1) are conventional. An FGD system is designated throwaway or recovery in terms of actual or potential marketing of the sulfur product in the U.S. Whether sludge to a pond or stabilized sludge for landfill, the SO_2 product from limestone or lime scrubbing is considered throwaway. The gypsumlike products, which might be made marketable in Japan or Europe, still are throwaway in the U.S. On the other hand, FGD systems producing sulfur, SO_2 , H_2SO_4 , or fertilizers are listed as recovery processes without regard to their location within the U.S.

The active ingredient in a nonregenerable sorbent is used only once for SO_2 sorption. In limestone and lime slurry processes, the sorbing slurry normally is recirculated over the SO_2 absorber but no regeneration of the limestone or lime chemical is involved. With a regenerable sorbent, the SO_2 -laden sorbent from the absorber is regenerated, usually by precipitating and filtering off its SO_2 component, before returning the sorbent to the absorber. In this type of system, the SO_2 equivalent of the loaded sorbent from the absorber is definitely greater (say 2-5 times) than that of the restored sorbent to the absorber. The third category of sorbent, that for an SO_2 derivative, is involved when the SO_2 in the flue gas is converted to another chemical, such as SO_3 or sulfur, before sorption.

3.2.2 *Process Listing*^{32,34,40,44,46,48-54}

Current FGD processes, worldwide, are listed in Table 3.2 according to the classification in Table 3.1. This listing omits SO_2 removal processes which have been associated with industrial applications only, such as copper (Cu) smelters, H_2SO_4 plants, and Claus plants. It also omits FGD processes which appear to be extinct, even though this distinction is not always clear cut. Some FGD processes have faded from lack of success or good prospects rather than definitive failure. Process promoters often are slow to admit this status unless they have taken up another process. However, no attempt has been made in Table 3.2 to compare and rate the competitive commercial versions of a given process.

Table 3.2. Current FGD Processes - World

Category; ^a SO ₂ Sorbent	Process Material	SO ₂ Product	Largest Unit (MW)	Process Name, Vendor, Sponsor ^b	Country
I. THROWAWAY PRODUCT					
A. NONREGENERABLE SORBENT					
1. Solution					
Alkaline seawater	Alkaline waste	To sea	10	Flakt (Norsk Hydro)	Norway
Na ₂ CO ₃	-	Na ₂ SO ₃ , NaHSO ₃ solution	3 @ 125	CEA	U.S.
2. Slurry					
CaCO ₃ (chalk)/ alkaline riverwater	-	CaSO ₄ to river	300	Bankside	England
CaCO ₃	-	CaSO ₃ /CaSO ₄ sludge	820	B&W, R-C, CE, PB, UOP, TVA	U.S.
CaCO ₃	-	CaSO ₃ /CaSO ₄ sludge	10	Bahco	Sweden
CaCO ₃	-	CaSO ₃ /CaSO ₄ cake	660	Mitsui-Chemico, Babcock-Hitachi, Fuji Kasui-Sumitomo, Chubu-MKK, IHI-TCA, Chemico-IHI, MHI	Japan
Ca(OH) ₂	-	CaSO ₃ /CaSO ₄ sludge	835	CEA, SCE, UOP, AAF, B&W, R-C, CE, Chemico	U.S.
Ca(OH) ₂	-	CaSO ₃ /CaSO ₄ sludge	10	Bahco	Sweden
Ca(OH) ₂	-	CaSO ₃ /CaSO ₄ sludge	145	Bischoff	W. Germany

(continued)

Table 3.2 (continued)

Category; ^a SO ₂ Sorbent	Process Material	SO ₂ Product	Largest Unit (MW)	Process Name, Vendor, Sponsor ^b	Country
Ca(OH) ₂	-	Gypsum cake	375	MHI, Nippon Steel, Mitsui-Chemico, Kobe Steel, Hitachi, Fuji Kasui-Sumitomo, IHI-TCA	Japan
3. Solid					
CaCO ₃	-	CaSO ₃ /CaSO ₄ sludge	400	CE (terminating)	U.S.
B. REGENERABLE SORBENT					
1. Solution					
1% H ₂ SO ₄	H ₂ O, CaCO ₃	Gypsum cake	23	Chiyoda 101	U.S.
1% H ₂ SO ₄	H ₂ O, CaCO ₃	Gypsum cake	350	Chiyoda 101	Japan
Sodium compounds	Na ₂ CO ₃ , Ca(OH) ₂	CaSO ₃ /CaSO ₄ /CaCO ₃ cake	40	General Motors	U.S.
Sodium compounds	Na ₂ CO ₃ , Ca(OH) ₂	CaSO ₃ /CaSO ₄ /CaCO ₃ cake	55;575 ^c	CEA/ADL, FMC, Envirotech	U.S.
CaCl ₂ /Ca(OH) ₂	Ca(OH) ₂	Gypsum cake	40	Saarberg-Holter	W. Germany
Na ₂ SO ₃	CaCO ₃ , H ₂ SO ₄	Gypsum cake	450	Kureha-Kawasaki, Showa-Denko	Japan
Al ₂ (SO ₄) ₃ •Al ₂ O ₃	CaCO ₃	Gypsum cake	30	Dowa Mining Company	Japan
Sodium acetate	Ca(OH) ₂	Gypsum cake	2	Kureha	Japan
3. Solid					
Activated carbon	H ₂ O, CaCO ₃	Gypsum cake	150	Hitachi	Japan

(continued)

Table 3.2 (continued)

Category; ^a SO ₂ Sorbent	Process Material	SO ₂ Product	Largest Unit (MW)	Process Name, Vendor, Sponsor ^b	Country
II. RECOVERABLE PRODUCT					
A. NONREGENERABLE SORBENT					
1. Solution					
Industrial application only, e.g., H ₂ SO ₄ plant					
B. REGENERABLE SORBENT					
1. Solution					
Sodium citrate	H ₂ S	Sulfur	0.5;60 ^c	BOM (citrate)	U.S.
Sodium citrate	Steam	SO ₂	1.5	Flakt-Boliden	Sweden
Na ₃ PO ₄	H ₂ S	Sulfur	0.05	Stauffer-Chemico	U.S.
Na ₂ SO ₃	Natural gas	Sulfur from SO ₂	115	W-L/AC	U.S.
Na ₂ SO ₃	Natural gas	H ₂ SO ₄ from SO ₂	220	Wellman-MKK	Japan
Na ₂ CO ₃	Coke, CO ₂	Sulfur from H ₂ S	100 ^c	AI	U.S.
Na ₂ CO ₃	Oil, CO ₂	Sulfur from H ₂ S	-	TSK	Japan
Na ₂ SO ₃	NaOH	Na ₂ SO ₃ crystal	-	Kurena	Japan
NH ₃ solution	Synthesis gas	Sulfur from SO ₂	2/30 ^d	Catalytic/IFP	U.S., France
K ₂ CO ₃	CO	Sulfur	7/0.35 ^d	CONSOL	U.S.
Glyoxylic acid	Steam	SO ₂	0.15	Spring-Nobel Hoechst	France
2. Slurry					
MgO	Heat	H ₂ SO ₄ from SO ₂	155	Chemico, UE	U.S.
FeS	Coal	Sulfur	1 ^c	SULF-X (PENSYS)	U.S.

(continued)

Table 3.2 (continued)

Category; ^a SO ₂ Sorbent	Process Material	SO ₂ Product	Largest Unit (MW)	Process Name, Vendor, Sponsor ^b	Country
3. Solid					
Activated carbon	Coal	Sulfur from SO ₂	20/47.5 ^d	FW/BF	U.S.
Activated carbon	Steam, H ₂ S from H ₂	Sulfur from H ₂ SO ₄	0.2	Westvaco	U.S.
Activated carbon	H ₂	Sulfur from SO ₂	42	BF	W. Germany
Activated carbon	H ₂ O	60% H ₂ SO ₄	60	Hitachi	Japan
CuO on alumina	H ₂	Sulfur from SO ₂	40, 0.6	Shell/UOP (SFGD)	Japan, U.S.
4. Melt					
Na ₂ CO ₃ /K ₂ CO ₃ /Li ₂ CO ₃	Coke	Sulfur	10	AI	-
C. SORBENT FOR SO ₂ DERIVATIVE (SO ₃)					
1. Solution					
H ₂ SO ₄	V ₂ O ₅ catalyst, H ₂ O	78% H ₂ SO ₄	110	Monsanto (Cat-Ox)	U.S.

^aSee Table 3.1.^bFor abbreviations, see Appendix B.^cUnder construction or planned.^dMW for absorption/MW for reduction to sulfur.

3.3 DESCRIPTIONS OF WORLD PROCESSES

3.3.1 Throwaway - Nonregenerable

Status

This is the main category of Table 3.1 for commercial FGD installations in the U.S. and a significant one for those in Japan and West Germany. In the U.S., this category includes all FGD units now in commercial operation and 90% of those under construction or currently planned for construction. Among the U.S. operating units, slurry-type sorbents are used at all locations, except at one site-specific station which uses a solution-type sorbent and at a second location which is converting from solid to slurry-type sorbent. With that conversion, both versions of limestone addition to the boiler (without and with scrubbers) will be extinct.

Solution Sorbents

The 10-MW Flakt-Hydro installation (Table 3.2) at the Porsgrunn Works of Norsk Hydro in Norway illustrates a site-specific FGD condition.⁴⁸ The process uses seawater as sorbent, and after neutralization and oxidation, the scrubber effluent is returned to the sea.

In this process (Fig. 3.2) ash is removed from boiler flue gas by multicyclones whose pressure drop is controlled by a gas bypass. After quenching with seawater, the flue gas flows upward through a saddle-packed absorber which is irrigated with seawater at about pH 8. Reaction between fresh seawater and SO_2 is thought to be from bicarbonate (HCO_3^-) and hydroxide ions (OH^-). Scrubbed flue gas leaves the absorber through a mist eliminator at about the temperature of the seawater, 50–55°F, and is reheated with a bypass stream of gas at about 700°F. SO_2 removal is about 90%. The scrubber effluent is neutralized and oxidized, before disposal to the sea, by mixing with the alkaline, oxygen-saturated effluent from a nearby MgO plant.

The commercial example of this category, the Combustion Equipment Associates' installation at Reid Gardner Station, Nevada Power Company, is described in Sec. 6. Its SO_2 sorbent is a solution of commercial-grade soda ash (Na_2CO_3).

FIG. 3.2. FLAKT - HYDRO PROCESS
 (THROWAWAY , NONREGENERABLE , SOLUTION)

*Slurry Sorbents*⁴²

A generalized limestone slurry process is shown in Fig. 3.3 -- more detailed treatment is given under commercial FGD. In this process limestone is wet ground and pumped to a slurry feed tank adjacent to the scrubbing area. This makeup slurry and recycled clear liquor from the disposal pond are fed to the absorber effluent hold tank for blending with the partially reacted slurry draining from the absorber. Slurry is recirculated between this hold tank and the absorber where it scrubs the flue gas. A bleed stream of reaction products is pumped from the hold tank to the waste disposal system or settling pond. The outlet gas from the scrubber passes through a slurry entrainment separator and is reheated before discharge to the stack. The slurry of reaction products in the waste disposal pond settles to form a CaSO_3 - CaSO_4 sludge containing about 40-50% by wt solids. Clarified pond water is returned to the scrubbing system to maintain closed-loop operation.

Lime slurry scrubbing is similar, in all areas except slurry preparation, to the limestone process described. To prepare the scrubbing slurry, lime is conveyed from a storage tank for slaking with recycled pond water. Lime slurry from the slakers is diluted with additional pond water and is pumped to a slurry feed tank for in-process storage. Slurry from this tank is blended with more pond water before feeding to the scrubber.

Waste product from either the limestone or lime system can be ponded as it comes from the scrubber or it can be thickened before disposal. When ponded directly, the pond provides settling of solids and clarification of liquor for recirculation. When thickened before disposal, the liquor for recycle is reclaimed before ponding. An additional sludge fixation step may be included to provide more stable and environmentally acceptable solids for disposal. Neither the technology nor costs for fixation are adequately defined.

Solid Sorbents^{25, 48, 55-61}

Limestone injection to the boiler is relatively simple requiring limestone receiving and storage facilities, a grinding installation, an injection system, and conveyors (Fig. 3.4). The injection equipment includes an air blower or compressor, transfer lines, and a set of nozzles. Tilttable

FIG. 3.3. LIMESTONE SLURRY PROCESS
(THROWAWAY, NONREGENERABLE, SLURRY)

FIG. 3.4. LIMESTONE INJECTION
(THROWAWAY, NONREGENERABLE, SOLID)

nozzles, which can vary the trajectory, are necessary to adjust for differences in limestone quality and in boiler operations. As an alternative to injection, the limestone may be ground with the coal and fed as a mixture through the coal burners.

Finely ground limestone is injected into the boiler and reacts with the SO_2 in the flue gas at the high temperatures in the boiler (1500–3000°F). It is commonly assumed that the reaction is twofold: (1) calcination of CaCO_3 to calcium oxide (CaO) and carbon dioxide (CO_2) and (2) sorption of SO_2 by CaO with oxidation to CaSO_4 . Oxygen is supplied from the excess air in the flue gas. The reacted particles move with the flue gas to the dust collecting equipment for removal. Dust removal and disposal systems must be enlarged to handle the increased quantity of particulate matter.

Removal of SO_2 by the limestone-boiler injection method is approximately 20–50% efficient. To increase removal capability, wet scrubbing was added downstream of the boiler as in a former installation at Hawthorn station of Kansas City Power and Light Company. This facility is described under Operational FGD in the U.S. (Sec. 6.2.2). All variations of limestone addition to the boiler are now obsolete.

3.3.2 Throwaway - Regenerable

*Solution Sorbents*⁶²

Cooled and saturated flue gas is contacted in an absorber tower with an aqueous solution containing a soluble Na compound, such as Na_2CO_3 or sodium hydroxide (NaOH) (Fig. 3.5). SO_2 in the flue gas is reacted with the scrubbing solution to form sodium sulfite (Na_2SO_3) and bisulfite (NaHSO_3). Some oxidation to sulfate occurs. Scrubber effluent is recycled from a hold tank beneath the scrubber to the precooler and the scrubber. Cleaned gas leaves the tower through a mist eliminator and is reheated before entering the stack.

A bleed stream of scrubber effluent is pumped to a stirred reactor tank where it is blended with slaked lime and slurried. CaSO_3 and CaSO_4 crystals are formed and the Na scrubbing solution is regenerated. The Ca crystal slurry then is pumped to a thickener to concentrate the slurry to about 30% solids. Na_2CO_3 is dissolved in water and added to the thickener to

FIG. 3.5. DOUBLE - ALKALI PROCESS
(THROWAWAY , REGENERABLE , SOLUTION)

replenish the Na lost in the wet cake product. Regenerated scrubbing solution leaves the thickener in the overflow and is returned to the absorber.

The concentrated Ca slurry is dewatered by vacuum filter and the cake is washed to decrease Na loss. Ponding or pit disposal can be used to contain the waste product cake. Filtrate and wash water from the filter are recirculated to the thickener.

Na makeup as soda ash can be added to the thickener overflow storage tank or to the effluent hold tank rather than to the thickener itself. Variations in sorbent include aluminum $\overline{Al_2(SO_4)_3}$ or ammonium sulfate $\overline{(NH_4)_2SO_4}$ and sodium acetate ($NaC_2H_3O_2$). Depending on the process, limestone may replace lime as the Ca source. A small purge stream of sorbing solution may be discarded or treated to control chloride and MgO buildup in the system.

American development of processes in this category usually follows the trend to Ca-sulfur throwaway products. Development in Japan has been toward the production of salable gypsum. This is done by adding an oxidation step after absorption to convert sulfite to sulfate before reaction with lime or limestone.

In the Saarberg-Holter process in West Germany,^{63,64} flue gas from the ESP is spray cooled, scrubbed, and demisted in a combination scrubber with a pressure drop of 10 in. water. The regenerable sorbent is a buffered solution of calcium chloride $\overline{Ca(OH)_2/HCl}$ which removes SO_2 by forming calcium bisulfite $\overline{Ca(HSO_3)_2}$, while changing from slightly alkaline to slightly acid. Makeup lime and oxidizing air are added to the scrubber effluent to precipitate gypsum and to restore the $CaCl_2$ and the pH of the sorbing solution. Gypsum is separated by a thickener and a vacuum filter and sold for industrial use. During 11,000 hr of operation, the 40-MW unit reportedly has had 90% availability (95% excluding downtime for planned modifications), 90+% SO_2 removal from 800 ppm SO_2 in flue gas, a power requirement of 1% (no reheat) of power plant capacity, a turndown to 40%, a closed-loop cycle, and no plugging.

The Chiyoda Thoroughbred 101 process⁶⁵ uses dilute H_2SO_4 as the absorbing solution. The unique design feature of the process is the combination of the absorbing and oxidation towers into one vessel. A high-quality gypsum is the sulfur product.

Solid Sorbents^{3,2}

The Hitachi carbon process (Fig. 3.6) requires flue gas relatively free of particulates to avoid fouling the carbon beds. ESPs remove flyash before the gas reaches the adsorption area. The adsorption tower has three fixed beds of activated carbon. Flue gas is directed to one bed where SO_2 is adsorbed on the carbon and oxidized in place to SO_3 . When adsorption is complete, gas flow is directed to another section while wash water flushes the first bed and absorbs the SO_3 . The resulting liquor is 20% H_2SO_4 . SO_2 removal is about 80% of the inlet SO_2 .

An aqueous slurry of pulverized limestone is added to the dilute acid in a stirred reaction tank. Large gypsum crystals are precipitated and are separated in batch centrifuges. Liquor from the centrifuges is recycled to the reaction tank; a purge stream is pumped from the centrifuges to a thickener. Clarified liquor from the thickener is neutralized with NaOH and discharged to the sea. Thickener underflow is returned to the limestone slurry tank. There is no recycle wash water because of the danger of plugging the carbon pores with gypsum, limestone, and soot particles.

3.3.3 Throwaway - SO_2 Derivative

No processes in this category are currently significant. The category is illustrated by a catalytic-oxidation type where the product H_2SO_4 is too dilute for commercial use and is neutralized with lime or limestone for a throwaway product.

3.3.4 Recoverable - Nonregenerable

NH_3 absorption of SO_2 was used in the fertilizer industry and for cleaning smelter gases before its consideration for FGD. Development of a process by Showa Denko KK (Tokyo)⁴⁰ was the first large-scale testing (25 MW) on fossil fuel stack gas. The test unit, scrubbing gas from an oil-fired process-steam plant, began operation in October 1966.

NH_3 is injected into the duct carrying the gas from the air heater (or ESP in the case of coal-fired boilers) to the scrubber. Early introduction of NH_3 helps reduce corrosion by SO_3 . Water is also injected into the duct to cool the gas. The cooled and partially treated gas flows into

FIG. 3.6. HITACHI CARBON PROCESS
(THROWAWAY, REGENERABLE, SOLID)

the absorber where it is contacted by the scrubbing liquor which contains ammonium sulfite $\overline{[(\text{NH}_4)_2\text{SO}_3 \cdot \text{H}_2\text{O}]}$ and bisulfite (NH_4HSO_3). Cleaned gas leaves the absorber through a mist eliminator. In the original scheme, the gas was not reheated before entering the stack. Work by TVA and others has shown that formation of a persistent fume is a problem with NH_3 scrubbing.

Absorber effluent drains into a surge tank where makeup NH_3 and water are added to convert NH_4HSO_3 to $(\text{NH}_4)_2\text{SO}_3 \cdot \text{H}_2\text{O}$. Scrubbing liquor is recycled from the surge tank to the absorber. A bleed stream of the liquor is pumped to an oxidizer where air is injected to convert the sulfite to $(\text{NH}_4)_2\text{SO}_4$. The $(\text{NH}_4)_2\text{SO}_4$ solution is pumped to a crystallizer for recovery of the sulfate product.

3.3.5 Recoverable - Regenerable

*Solution Sorbents*⁴²

The Wellman-Lord (W-L) scrubbing process coupled with the Allied Chemical (AC) process for SO_2 reduction to sulfur illustrates this category. A detailed flow diagram of this process is shown in Sec. 5.2.3. Particulates in the flue gas are removed in a venturi scrubber, neutralized with slaked lime, and pumped to the pond. Pond water is recycled and makeup water is added to replace pond losses. Downstream from the particulate scrubber, flue gas is contacted with an Na solution in a valve-tray absorber to remove the SO_2 . Cleaned gas leaves the absorber through a mist eliminator and is reheated before entering the stack.

A part of the absorber effluent stream is cooled for crystallization of sodium sulfate (Na_2SO_4) which is separated in a centrifuge. Its formation increases raw material makeup. After drying in a rotary dryer, the Na_2SO_4 is conveyed to a product storage area -- disposal can be a problem in many locations.

The liquor remaining after Na_2SO_4 removal and the remaining absorber effluent are pumped to a regeneration area where an evaporator-crystallizer produces Na_2SO_3 crystals and SO_2 gases. The Na_2SO_3 is returned to a dissolving tank where soda ash is added to make up for handling losses and for Na lost in Na_2SO_4 formation. An antioxidant also is added to minimize the oxidation of sulfite to sulfate in the absorber.

Regenerated SO_2 in the off-gas is cooled and compressed to remove moisture. It is then fed to a reduction unit where it is mixed with natural gas and the gas combination is preheated by indirect heat exchange with the hot reduced gas. Reduction to this reduced gas, comprising sulfur, hydrogen sulfide (H_2S), CO_2 , and water, occurs in the reduction reactor system. After cooling, the reduced gas mixture enters a condenser-converter-condenser system where most of the H_2S and SO_2 are converted to sulfur. The molten sulfur is collected in a receiving pit and is held in steam-heated tanks for shipment. Tail gas from the reduction unit is incinerated with air and natural gas to oxidize any remaining H_2S to SO_2 and the off-gas from incineration is returned to the scrubbing system. Reduction systems based on coal gasification now are being evaluated as alternatives to the above natural gas-based type.

A second illustration of this category with a difference in approach to absorption is the Atomics International (AI) aqueous carbonate process (Fig. 3.7). Cyclones lower the ash content of the flue gas to 1.0 gr/sft^3 or less. Flue gas then flows through a spray dryer-scrubber where it is contacted with atomized droplets of Na_2CO_3 solution. The Na_2CO_3 reacts with the SO_2 forming Na_2SO_3 and Na_2SO_4 . The heat of the flue gas evaporates the water and the particulate matter (Na_2SO_3 , Na_2SO_4 , unreacted Na_2CO_3 , and ash) is carried by the flue gas to the collection system. Cyclones followed by an ESP clean the gas which is then reheated and vented to the stack.

The dry product is conveyed to a reducer where air and coke are introduced to melt the product and reduce the sulfite and sulfate to sodium sulfide (Na_2S). The melt is quenched and dissolved in water. Insolubles, consisting mainly of ash and carbon, are filtered out. The Na_2S -rich solution is pumped to the carbonation area where CO_2 -rich off-gas from the reducer produces regenerated Na_2CO_3 solution and H_2S which is used in the Claus production of sulfur.

Slurry Sorbent^{4,2}

Several MgO scrubbing processes have been developed in various countries around the world. The description given here will follow generally the Chemico-Basic system (Sec. 5.2.1). Flue gas is contacted in a venturi scrubber to remove particulates from the gas with a slurry of recycled pond

FIG. 3.7. AQUEOUS CARBONATE PROCESS - ATOMICS INTERNATIONAL
(RECOVERABLE, REGENERABLE, SOLUTION)

water and flyash. Scrubber effluent is neutralized with slaked lime as required and pumped to a disposal pond. A second venturi is used to contact the SO_2 in the flue gas with MgO slurry. Cleaned gas from this scrubber is reheated before entering the stack.

Effluent from the SO_2 absorber is passed through wet screens for thickening. The magnesium sulfite hexahydrate ($\text{MgSO}_3 \cdot 6\text{H}_2\text{O}$) in the thickened slurry is thermally converted to magnesium sulfite trihydrate ($\text{MgSO}_3 \cdot 3\text{H}_2\text{O}$) and pumped to centrifuges for separation. Centrifuge centrate and underflow from the wet screens are returned to the slurry preparation area and the SO_2 absorber loop. The centrifuge cake is dried in an oil-fired fluidized bed dryer. Dryer off-gas is cleaned in a cyclone and bag filter. The MgSO_3 solids from dryer, cyclone, and bag filter are fed to a calciner where, in the presence of coke, MgO and SO_2 are generated. The regenerated MgO is returned to the slurry preparation area. The SO_2 off-gas from the calciner is cleaned, cooled, and mixed with air before entering a conventional plant for the production of 98% H_2SO_4 .

Solid Sorbent⁵³

A moving-bed reactor with transverse gas flow is used in the Bergbau-Forschung (BF) process to contact flue gas with carbon adsorbent (Fig. 3.8). Beds of activated char are placed in a silo-type adsorber tower. Flue gas flows from an ESP into the center of the adsorber and radially outward through the beds. Regenerated char, added continuously at the top of the tower, moves downward and loaded char is withdrawn continuously at the bottom and conveyed to the regeneration unit. Cleaned gas flows from the adsorber to the stack.

Carbon fines and the flyash collected in the adsorber are screened from the loaded char prior to regeneration in a moving-bed reactor. Sand is heated to 1200-1300°F and pneumatically conveyed by the combustion gases to the top of the regenerator where it is mixed directly with the loaded char. The mix at a temperature of 900-1200°F moves through the reactor and is discharged to a vibrating screen deck to separate the char and sand. Regenerated char is recycled to the adsorber and the sand is reheated and recycled to the regenerator.

FIG. 3.8. BERGBAU-FORSCHUNG FGD PROCESS
(RECOVERABLE, REGENERABLE, SOLID)

The gas emitted from the regenerator is rich in SO_2 and is drawn off to a Claus unit. Low-Btu gas is used to reduce a portion of the SO_2 to H_2S , which is reacted with the remaining SO_2 to produce sulfur. The sulfur is condensed and is stored in liquid form in a heated tank. Any unconverted H_2S is incinerated to SO_2 and returned to the adsorber.

The Shell FGD process licensed by Universal Oil Products (UOP) uses copper oxide (CuO) on alumina grids to react with SO_2 . The copper sulfate (CuSO_4) formed by the reaction is regenerated and SO_2 is evolved by passing a reducing gas over the laden grids. The end product may be SO_2 , H_2SO_4 , or sulfur, depending upon market conditions.

Melt Sorbent⁵¹

The AI molten carbonate process employs a molten mixture of Na, potassium (K), and lithium carbonates to react with the SO_2 in the flue gas. The process operates at about 850°F . In the scrubber the sulfur oxides (SO_x) react with the melt to produce sulfites and sulfates. CO_2 is evolved. The cleaned gas leaves the scrubber through a wire mesh demister. In a reducer the sulfites and sulfates are reduced to sulfides with petroleum coke. More CO_2 is evolved and the sulfides are reacted in a regenerator with a gaseous mixture of this CO_2 and of steam. The H_2S produced in the regenerator is sent to a Claus plant for conversion to sulfur. Regenerated molten carbonates flow into a melt tank for recycle to the scrubber. Carbonate makeup is added to the regenerated carbonates at the melt tank. Work on this process is no longer active.

3.3.6 Recoverable - SO_2 Derivative

Solution^{42, 53, 66-68}

The Monsanto Cat-Ox process (Fig. 3.9) requires a gas temperature in the range of $850\text{--}900^\circ\text{F}$ for the catalytic vanadium pentoxide (V_2O_5) oxidation of SO_2 to SO_3 . With this requirement, the SO_2 converter is located upstream of the economizers and air heaters, in a new unit design. In retrofit installations, preheaters are used to achieve the required gas temperature. High-efficiency ESPs are necessary to reduce the particulate

FIG. 3.9. MONSANTO CAT-OX PROCESS
(RECOVERABLE, SORBENT FOR SO₂ DERIVATIVE, SOLUTION)

loading of the gas to 0.005 gr/sft^3 or less to minimize fouling of the catalyst beds in the converters. From the converter the gas flows through the economizers and air heaters of the power plant unit before entering the absorber. In retrofit installations, additional heat exchangers are included to recover the heat from the gas stream.

SO_3 and water vapor in the gas stream are absorbed in packed towers using a recycled stream of H_2SO_4 from the acid coolers. Cleaned gas leaves the absorber through high-efficiency Brink fiber demisters to remove entrained acid and water droplets. Reheating is not necessary because the gas leaves the absorber at about 240°F .

To reclaim energy from the system, heat recovery loops are added to heat boiler feed water for the steam plant and to preheat air to the furnace air heater. The acid effluent from the absorber flows through two circulation coolers to a final-product acid cooler before storage and shipping. Acid produced is about 80% H_2SO_4 . Now inactive, a large-scale demonstration of the Cat-Ox process is described in Sec. 5.2.2.

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4 FGD DEVELOPMENT OUTSIDE THE U.S.

4.1 JAPAN^{42,69,70}

Because of physical congestion, Japan has experienced very high pollution from its utilities and heavy industry. By about 1970 it had undertaken heavy development programs for FGD and these led to faster commercialization than occurred in the U.S. Currently, Japan has about 12,000 MW (Table 4.1) of utility capacity equipped with FGD, in addition to its installations on smelters, industrial boilers, etc.

Several FGD factors in Japan differ from those in the U.S. and a somewhat different profile of FGD processes has emerged. As a result, some processes are not as directly transferable to the U.S. as might be inferred from their basic features. Moderate-sulfur oil is the most common fuel for boiler and reheat purposes and gypsum of wallboard quality is the favored product. This involves an emphasis on oxidation to gypsum which generally has not been justified in the U.S. Also, Japan has not had a primary requirement for closed-loop operation.

Table 4.1. Major FGD Installations at Power Plants Outside the U.S.

Category Sorbent	Japan	England	W. Germany	Total
	(MW)			
Throwaway - nonregenerable				
CaCO ₃ slurry	5,920	300	-	6,220
Ca(OH) ₂ slurry	2,180	-	145	2,325
Throwaway - regenerable				
CaCl ₂ solution	-	-	-	40
H ₂ SO ₄ solution	850	-	-	850
Na solution	2,900	-	-	2,900
Recovery - regenerable				
Na solution	345	-	-	345
Activated carbon	210	-	40	250
Total	12,405	300	225	12,930

Limestone, lime, and double-alkali processes are used in most commercial operations in power plants. The Chiyoda 101 process has several installations with good reputations and its 23-MW prototype unit at Gulf Power Company in Florida has been tested for U.S. application on coal. Among recovery processes, only the W-L process has been successful.

4.2 EUROPE^{42,71}

England's single FGD installation was described in Sec. 3.1.1. Since the present policy in England is for tall-stack dispersion, FGD development is inactive.

West Germany emphasized FGD development during the late 1960's and early 1970's but only its Bischoff lime process has reached commercial importance. Now the Saarberg-Holter process which has many potential advantages is being tested at the 40-MW prototype level. One activated carbon installation (40 MW) continues.

In spite of industrial concentrations, West Germany began strong enforcement only recently. A number of new power plant units are under construction or planned and they appear to be evaluating FGD processes from the U.S. and Japan as well as from native development. Coal is a common fuel in West Germany and disposal options often are more restrictive than in the U.S. Since gypsum is marketable to the European building industry, processes with gypsum capability can have greater attraction than in the U.S.

At present, West Germany has only nominal FGD (225 MW in Table 4.1) and other European countries have none on regular power plant installations. France is attempting to market its IFP (Institut Francais du Petrole) NH_3 process, part of which has been tested on a 30-MW slipstream of an oil-fired power plant.

5 NONCOMMERCIAL FGD IN THE U.S.

5.1 FLUE GAS DEVELOPMENT UNITS IN THE U.S.^{43,48-50,72,73}

The currently significant FGD development units in the U.S. are grouped in Tables 5.1, 5.2, and 5.3 according to size. The demonstration-sized category in Table 5.1 has a minimum size of 50 MW but unit sizes range from 50-277 MW. (Demonstration-sized units sometimes are defined at 100 MW, minimum.)³² The prototype size, defined here as 10-50 MW, has units in Table 5.2 at 10-23 MW. Pilot plants (Table 5.3) are nominally under 10 MW and many units are at 1 MW or less. All units listed as developmental were built for research, development, or demonstration, and in most cases, their funding has been shared by federal agencies (especially EPA), utilities, process vendors, and equipment suppliers. They are not intended as a means to full compliance with air quality control standards but some demonstration-sized units could evolve into that position.

Louisville Gas and Electric's 65-MW scrubbing installation at Paddy's Run No. 6 is listed as a demonstration rather than a commercial unit because of its size, its intermittent operation (peaking), and its special carbide-lime sorbent.

5.2 PROCESS DEVELOPMENTS

5.2.1 MgO Slurry Process^{42,50,53,74-76}

MgO scrubbing with H₂SO₄ production (Fig. 5.1) has been installed on demonstration bases at three sites:

Utility	Station	FGD (MW)	Vendor
Boston Edison	Mystic	150	Chemico
Potomac Electric Power	Dickerson	95	Chemico
Philadelphia Electric	Eddystone	120	United Engineers/ Philadelphia Electric

A 150-MW MgO scrubbing system operated intermittently from April 1972 through June 1974 at the oil-fired Mystic 6 station of Boston Edison Company. The facilities for MgO regeneration and for H₂SO₄ production were located

Table 5.1. Principal Demonstration-Scale FGD Development Units in the U.S.

Item No.	Power Plant			FGD Installation								Fuel			
	Utility	Station	Boiler No.	Total MW	Startup Date	Vendor ^a	Process	New/Retrofit	Throw-away	Bypass	Total Trains	Type	Btu/lb	% S	% ash
1	Boston Edison	Mystic	6	150	4/72-6/74	Chemico	MgO scrubbing	R	No	Yes	1	#6 fuel oil	18,500	2.0	0.07
2	Illinois Power	Wood River	4	110	9/72-7/73	MN	Cat-Ox	R	No	Yes	1	Coal	10,800	2.9-3.2	10
3	Louisville Gas and Electric	Paddy's Run	6	65	4/73	CE	Carbide lime scrubbing	R	Yes	No	2	Coal	11,500	3.5-4.0	13
4	Potomac Electric and Power	Dickerson	3	95	9/73-8/75	Chemico	MgO scrubbing	R	No	No	2	Coal	11,700	2.0	14
5	Public Service of Colorado	Valmont	5	50	10/74-1/76	UOP/PSCC	Limestone scrubbing	R	Yes	-	1	Coal	10,780	1.0	6-7
6	Philadelphia Electric	Eddystone	1A	120	9/75	UE/PECO	MgO scrubbing	R	No	Yes	1	Coal	12,100	2.3	12
7	Arizona Public Service	Four Corners	5A	160	2/76-12/76	SCE	Lime scrubbing	R	Yes	-	1	Coal	-	0.7	-
8	Detroit Edison	St. Clair	6	163	5/76, 1/77	PB	Limestone scrubbing	R	Yes	Yes	2	Coal	9,550	0.3-0.4	4
9	North Indiana Public Service	D. H. Mitchell	11	115	(8/76) 5/77	DP/AC	Na solution/SO ₂ reduction	R	No	-	-	Coal	-	3.2	-
10	Niagara Mohawk Power Corporation	Charles R. Huntley	6	100	6/78	AI	Aqueous carbonate	R	No	-	-	Coal	-	2.5-4.5	-
11	Louisville Gas and Electric	Cane Run	6	277	2/79	CEA/ADL	Double alkali	R	Yes	-	-	Coal	-	0.5-0.8	-
12	St. Joe Minerals	George F. Weaton	1	60	0/80	BM	Citrate	R	No	Yes	1	Coal	12,500	2.5	<15

(continued)

Table 5.1 (continued)

Item No.	SIP ^b	Pollutant Control						FGD Product		Flue Gas Reheat		
		Particulate			SO ₂			Material	Disposal	Mode	Heat Source	Temp. to Stack, °F
		Device	Make ^a	%	Device	Make ^a	%					
1	0.55 ³	Venturi scrubber	Chemico	55	Venturi scrubber	Chemico	90	SO ₂ for H ₂ SO ₄	Market	None	None	-
2	1.80 ¹	ESP	R-C	99.9	Catalyst bed	MN	85	78% H ₂ SO ₄	Market	None	None	250-270
3	1.20 ¹	ESP	-	99	Two-stage marble bed	CE	80	Unstabilized sludge	Unlined pond	Direct	Natural gas	166-176
4	1.00 ⁶	ESP, variable-throat venturi	R-C, Chemico	-	Fixed-venturi scrubber	Chemico	90	SO ₂ for H ₂ SO ₄	Market	-	-	-
5	- ¹⁰	TCA	UOP	95	TCA	UOP	45-85	Unstabilized sludge	Lined pond	-	-	-
6	0.60 ²	ESF, single stage adjustable venturi rod	JW;PB; EV	99.9	Two-stage adjustable venturi rod	EV	90	SO ₂ for H ₂ SO ₄	Market	Direct	Fuel oil	237
7	20M ¹⁰	ESP	-	-	Four-stage weir horizontal scrubber	-	-	-	-	-	-	-
8	1.50 ⁶	ESP	-	99.7	Variable-throat venturi scrubber	PL	90	Unstabilized sludge	Lined pond	Direct	Oil	250
9	1.20 ¹	-	-	-	-	-	-	Sulfur	Market	-	-	-
10	1.90 ³	ESP	-	-	Spray dryer	AI	90	Sulfur	Market	None	None	175
11	1.20 ¹	ESF	-	-	-	-	95	-	-	-	-	-
12	1.20 ²	ESE	-	-	Packed tower	-	90	Sulfur	-	Indirect	Steam	166

^aSee Appendix B.^bSee Appendix C.

Table 5.2. Principal Prototype FGD Development Units in the U.S.

Item No.	Power Plant			Total MW	Startup Date	FGD Installation					Total Trains	Fuel			
	Utility	Station	Boiler No.			Vendor ^a	Process	New/Retrofit	Throw-away	Bypass		Type	Btu/lb	% S	% ash
1	Metropolitan Edison	Portland	2	15	10/67-10/68	EN	Catalytic oxidation	R	No	-	-	-	-	-	-
2	Tennessee Valley Authority	Shawnee	10A	10	4/72	UCP	Lime/limestone scrubbing	R	Yes	No	1	Coal	12,000	2.9-3.7	13-16
3	Tennessee Valley Authority	Shawnee	10B	10	4/72	Chemico	Lime/limestone scrubbing	R	Yes	No	1	Coal	12,000	2.9-3.7	13-16
4	Tennessee Valley Authority	Shawnee	10C	10	4/72-7/73	TVA	Lime/limestone scrubbing	R	Yes	No	1	Coal	12,000	2.9-3.7	13-16
5	Consolidated Edison	Arthur Kill	20	10	7/73-4/75	AI	Molten carbonate	R	No	-	-	#6 fuel oil	18,500	0.3	0.015
6	Gulf Power Company	Scholz	1A	20	2/75-7/76	CEA/ADL	Double alkali	R	Yes	Yes	1	Coal	11,500-14,100	0.9-3.1	10-14
7	Gulf Power Company	Scholz	1B-2B	23	3/75-5/76	CS	Chiyoda 101	R	Yes	Yes	1	Coal	11,500-14,100	0.9-3.1	10-14
8	Gulf Power Company	Scholz	2A	20	2/76-5/76	FW	Activated carbon	R	No	Yes	1	Coal	13,100	1.5-3.1	12

(continued)

Table 5.2 (continued)

Item No.	SI ^{2b}	Pollutant Control						FGD Product		Flue Gas Reheat		
		Particulate			SO ₂			Material	Disposal	Mode	Heat Source	Temp. to Stack, °F
		Device	Make ^a	%	Device	Make ^a	%					
1	-	ESP	-	99	Catalyst bed	MN	-	H ₂ SO ₄	Market	None	None	250-270
2	-	ESP, three-stage TCA	UOP	99.5	Three-stage TCA	UOP	85	Stabilized/unstabilized sludge	Lined/unlined pond	Indirect	Hot air	225
3	-	ESP, variable-throat venturi	Chemico	99.5	Spray tower	-	80	Stabilized/unstabilized sludge	Lined/unlined pond	Indirect	Hot air	225
4	-	ESP	CE	99.5	Marble bed	CE	-	Stabilized/unstabilized sludge	Lined/unlined pond	Indirect	Hot air	225
5	-	ESP	-	99	Spray chamber	AI	90	Sulfur	Market	Direct	#6 fuel oil	870
6	-	ESP, variable-throat venturi	BU, CEA/ADL	99.5	Tray tower, spray tower or deentrainment separator	CEA/ADL	99	Unstabilized sludge	Clay-lined pond	Direct	#2 fuel oil	200
7	-	ESP, venturi scrubber	BU, CY	99.5	Packed tower	CY	95	Gypsum (CaSO ₄ •2H ₂ O)	Lined pond	Direct	#2 fuel oil	200
8	-	ESP	BU	99.5	Adsorption columns	FW	96	Sulfur	Market	None	None	328

^aSee Appendix E.^bSee Appendix C.

Table 5.3. Principal Pilot-Plant FGD Development Units in the U.S.

Item No.	Power Plant		FGD Installation						Fuel			Particulate Removal		
	Utility	Station	Boiler No.	FGD MW	FGD Startup	FGD Vendor ^a	Process	Throw-away	Vol % SO ₂ , Inlet Flue Gas	Type	% S	Device	Make ^a	%
1	Magma Copper Company Smelter	San Manuel, Arizona	-	0.2	10/70-4/71	EM	Citrate	No	1.0-1.5	-	-	-	-	-
2	Tennessee Valley Authority	Colbert	3-4	1-3	-	TVA	Lime/limestone scrubbing	Yes	0.2-0.28	Coal	3.0-4.0	Venturi scrubber	TVA	99
3	Northern States Power	Black Dog	1	6.5	3/73-5/74	CE	Limestone scrubbing	Yes	0.08	Coal	0.8	Venturi-rod scrubber	CE	99
4	Pennsylvania Power and Light	Sunbury	4	5.0	9/73	JW	Limestone scrubbing	Yes	0.2-0.24	Coal	1.6-5.5	ESP, vertical spray chamber	JW	93
5	Stauffer-Chemico	Norwalk	-	0.05	1/74-6/74	Chemico	Na phosphate scrubbing	No	0.07-0.28	Oil	-	-	-	-
6	Bunker Hill Co. lead smelter	Kellogg, Idaho	-	0.5	2/74-5/74	BM	Citrate	No	0.3-1.0	-	-	Baghouse	-	-
7	Pfizer, Inc.	Terra Haute, Indiana	-	1.0	3/73-9/74	PB	Citrex	No	0.1-0.2	-	-	Venturi scrubber	-	-
8	Texas Utilities	Big Brown	-	-	-	Chemico	Lime scrubbing	Yes	-	-	-	-	-	-
9	Texas Utilities	Big Brown	-	-	-	MN	Double-alkali "Calsox"	Yes	-	-	-	-	-	-
10	Texas Utilities	Big Brown	-	-	-	R-C	Limestone scrubbing	Yes	-	-	-	-	-	-
11	Texas Utilities	Monticello	-	-	-	Chemico	Limestone scrubbing	Yes	-	-	-	-	-	-

(continued)

Table 5.3 (continued)

Item No.	SO ₂ Removal			FGD Product		Comments
	Device	Make ^a	%	Material	Disposal	
1	-	-	93-99	-	-	This preliminary citrate pilot plant demonstrated that the citrate process was capable of substantial SO ₂ removal.
2	Two-stage TCA	TVA	85	Stabilized/ unstabilized sludge	Lined/ unlined pond	This facility, jointly operated and funded by EPA, EPRI, and TVA Power Research Staff, has run extensive research programs in lime/limestone wet scrubbing. These programs included operating a venturi and TCA in series and testing a horizontal scrubber. Reheat, corrosion, and sludge disposal studies were conducted. The latest program is the operation of a cocurrent downdraft scrubber.
3	Single-stage marble bed	CE	50	-	-	Black Dog was retrofitted with the Chemico scrubbing system to gather information for the Sherburne County station. Several basic changes were made.
4	Vertical spray chamber	JW	85	Stabilized sludge	Clay-lined pond	The unit has demonstrated that with a high L/G ratio, good SO ₂ removal (85%) can be obtained with near stoichiometric addition of limestone.
5	Packed tower absorber	-	98-99	Sulfur	Market	Coal-burning operation was simulated by adding flyash directly to the flue gas. Unit was terminated 6/74.
6	Packed wet scrubber	-	96-99	Sulfur	Market	
7	Tray absorber	PB	95-97	Sulfur	Market	Sulfur was produced at 99.96% purity.
8	-	-	-	-	-	Texas Utilities' first effort in FGD research was to test the Chemico lime scrubbing system.
9	-	-	-	-	-	The "Calsox" process was said to be acceptable, but it was concluded that there was not enough full-scale experience at that time.
10	-	-	-	-	-	This process has been working well at Arizona Public Service Cholla Station. Research-Cottrell was awarded contracts for Martin Lake 1, 2, 3, and 4.
11	-	-	-	-	-	Texas utilities set up this plant to test Chemico's new spray scrubber with limestone.

^aSee Appendix A.

FIG. 5.1. MAGNESIA SLURRY - REGENERATION PROCESS FLOW DIAGRAM

remotely, at an existing H_2SO_4 plant in Rhode Island. This FGD project was installed as a research prototype and its operation and maintenance varied widely during the test program. During the first year of operation, scrubber availability was only 17% and maintenance was extremely heavy, but by March-June 1974, it had improved to an average of 76% with 1 mo at 87%. The longest continuous period of scrubber operation was 7 days. Throughout the program, only minor scaling of the absorber and no plugging of the mist eliminator occurred. Also, there was no plugging of pumps or piping systems after the early slaking problems had been solved. MgO losses at the scrubber and calciner remained excessive (about 10% at the end of the program) and they prevented an evaluation of impurities buildup or of the SO_2 reactivity of continuously recycled sorbent. The SO_2 removal efficiency was generally above 80% and during a test period in 1974 it averaged 91% while using regenerated MgO .

The 95-MW MgO scrubbing system at the coal-fired Dickerson station of Potomac Electric Power Company (PEPCO) was operated intermittently from September 1973 to August 1975. Calcining operations, delayed until July 1974, were at the Rhode Island H_2SO_4 plant which previously had served the Boston Edison test program. Due to mismatches in scrubbing and calcining schedules, the scrubber operated with long periods of virgin, then of regenerated, MgO . The different wetting and agglomerating properties of the two varieties led to persistent problems in slaking and injection. Also at the Dickerson station, the MgSO_3 was predominantly hexahydrate whereas at the Mystic station it had been mainly trihydrate. Handling difficulties resulted. No significant scaling problems occurred in the system but erosion and corrosion of carbon steel pumps and piping were severe. Particulate removal was consistently efficient, providing an outlet loading from the scrubber below 0.02 gr/sft^3 ; SO_2 removal was consistently above 90% when operating at design conditions.

The SO_2 scrubber at Philadelphia Electric's Eddystone Station was started in October 1975 and the offsite MgO regeneration facility first processed MgSO_3 the last part of that month. The SO_2 scrubber was shut down when the Olin Chemical Plant at Paulsboro, New Jersey, an H_2SO_4 producer and prime contractor for regeneration, permanently closed. SO_2 removal operations were restarted in May 1977; relocation of the MgO regeneration facilities are

complete. The installation of SO₂ scrubbers on the balance of unit 1 and the design of an FGD system for unit 2 will depend on the performance of the existing 120-MW system on unit 1.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are given as: MgO makeup, 500 lb/hr; No. 6 fuel oil for drying and calcining, 130 MBtu/hr; high-pressure steam for reheating stack gas to 175°F, 77 MBtu/hr; low-Btu gas for sulfur production (not required for H₂SO₄ production), 145 MBtu/hr; coke for reduction of magnesium sulfate (MgSO₄), 5 MBtu/hr; and electricity for fans and pumps, 9 MW.⁵³

5.2.2 Catalytic Oxidation Process^{42,53,66-68}

Catalytic oxidation has been installed at two sites:

Utility	Station	FGD (MW)	Vendor
Metropolitan Edison	Portland	15	Monsanto
Illinois Power	Wood River	110	Monsanto

Design efficiencies for SO₂ removal are 85% or greater; 80% H₂SO₄ is produced.

Favorable data from the integrated 15-MW Cat-Ox installation at Portland Station, Metropolitan Edison, led to a 110-MW retrofit demonstration at Wood River unit 4 of Illinois Power Company. Because efficient conversion of SO₂ to SO₃ requires gas temperatures of 850-900°F, flue gas preheat before the converter is necessary in retrofit design. Natural gas, and then fuel oil, was used for the inline preheat burners at Wood River. This caused excessive contamination of the catalyst and the preheat system was redesigned for an external oil-fired burner. In addition to catalyst fouling, mechanical and structural problems were severe. After sporadic operation, shutdown occurred in mid-1975. Funding has expired and no plans for rebuilding have been announced.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are given as: catalyst makeup, 15-20 lb/hr and electricity for precipitators and fans, 13 MW.⁵³ If a hot ESP were used, as in a new installation, no gas reheating would be required and a slight net production of energy would result.⁷⁶

5.2.3 Na Solution - SO_2 Reduction Process^{42,50,77,78}

A W-L Na_2SO_3 solution scrubbing unit (Fig. 5.2) has been installed at the D. H. Mitchell Station of Northern Indiana Public Service (NIPSCO). Startup was in May 1977. This is the first large-scale Na process unit in the U.S. (115 MW) and the first full-scale demonstration of the process using flue gas from high-sulfur (3.5%) coal. SO_2 removal of 90% or higher has been achieved on oil-fired boilers in Japan with the capability of rapid turndown to 35%. Sulfite oxidation is a process deficiency which complicates its general use in the U.S.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are estimated as: soda ash, 2,120 lb/hr; high-pressure steam for reheating stack gas to 175°F, 77 MBtu/hr; low-pressure steam for evaporation, 170 MBtu/hr; low-Btu gas for reduction, 92 MBtu/hr (natural gas is used as reductant at NIPSCO); and electricity for blowers and pumps, 12 MW.⁵³

5.2.4 Aqueous Carbonate Process^{53,79}

An Al aqueous carbonate system is being planned, at a size to be determined, for the Charles R. Huntley Station of Niagara Mohawk Power Corporation. The process is designed to remove 90-95% of the SO_2 from the flue gas and sulfur will be produced.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are estimated as: petroleum coke for reduction of Na_2SO_3 and for heat for the molten salt reducer, 13,000 lb/hr; No. 6 fuel oil for the Claus plant incinerator, 32 MBtu/hr; and electricity for blowers and pumps, 11 MW.⁵³

5.2.5 Double-Alkali Process^{61,62,80}

The full 277-MW gas stream at Louisville Gas and Electric's Cane Run No. 6 is being fitted with Combustion Equipment Associates' double-alkali system. The system is designed for 95% SO_2 removal. A disposable filter cake (principally CaSO_3) of at least 55% solids will be produced.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are estimated as: lime, 19,500 lb/hr; soda ash, 800 lb/hr; high-pressure steam for reheating stack gas to 175°F, 77 MBtu/hr; and

electricity, 3.6 MW.⁵³ In the prototype system at Scholz Steam Plant, Gulf Power Company, the greatest source of difficulty has been filter performance but improved materials of construction and modified design have improved this area.

5.2.6 Citrate Process^{50,53}

The citrate process of BOM will scrub an equivalent 60 MW at George F. Weaton Station, St. Joe Minerals. The process is expected to have more than 90% SO₂ removal. Sulfur will be the end product.

Major requirements for a 500-MW system burning coal at 3.5% sulfur and 12,000 Btu/lb are estimated as: soda ash, 1,100 lb/hr; citric acid makeup, 30 lb/hr; high-pressure steam for reheating stack gas to 175°F, 77 MBtu/hr; low-pressure steam for sulfur melting and Na₂SO₄ production, 40 MBtu/hr; low-Btu gas for H₂S production, 103 MBtu/hr; and electricity for fans and pumps, 9 MW.⁵³ The method of H₂S production -- reaction of product sulfur with reducing gas from coal gasification -- has not yet been demonstrated. Scale-up of the entire FGD unit is from a nonintegrated pilot plant of about 0.5 MW.

5.2.7 Other^{50,53}

A 20-MW prototype Chiyoda 101 system has been operated successfully at Scholz Steam Plant. Removal efficiency is 90%. The gypsum end product has potential use in the wallboard industry. There are 15 commercial installations of this process in Japan treating gas from oil-fired utility boilers, Claus sulfur plants, and industrial waste incinerators.

The Foster Wheeler - Bergbau-Forschung (FW-BF) dry adsorption (activated carbon) process also was installed as a 20-MW prototype unit at Scholz Steam Plant. (Up to 99% SO₂ removal has been achieved by BF in West Germany.) The SO₂ removal section had brief and troubled operation. The unit for reducing SO₂ to sulfur, using anthracite coal, did not reach regular operation and the entire system is inactive.

An AI molten carbonate process has been installed on a 10-MW equivalent gas stream at the Arthur Kill Power Plant of Consolidated Edison. Removal of SO₂ is 90% efficient. H₂S from the process is converted to sulfur in a Claus

plant. The process operates at 850°F and all process equipment and piping are heated to prevent solidification of the melt. The scrubber was operated in the summer of 1973 but the regeneration system was never operated.

New processes continue to be proposed. For instance, in the SULF-X process^{81,82} a suspension of ferrous sulfide (FeS) has been used to remove SO₂ and NO_x. The FeS reacts with SO₂ to produce FeS₂ and ferrous oxide (FeO). The reaction products are heated (with coal and air) to regenerate FeS and produce sulfur. The reduction of SO₂ therefore occurs in the absorber and avoids a separate and expensive stage for SO₂ reduction. This process is now entering the pilot-plant phase.⁸³

Absorption by dry sodium bicarbonate (NaHCO₃) was tested several years ago by Southern California Edison. Commercial NaHCO₃ or nahcolite,⁸⁴ a natural form of NaHCO₃, was injected into the flue gas stream at Southern California's Alamos Station. Both forms of bicarbonate gave 70% SO₂ removal at a stoichiometry of 1 and 90% removal at a stoichiometry of 2. Since that time, Wheelabrator-Frye and Air Products Company have tested the process at various flue gas temperatures. Absorption increases with temperature increase; at approximately 500°F and 1.35 stoichiometry, SO₂ removal was 95%. Superior Oil has tested the process using a moving-bed reactor. Removal of 84% using commercial bicarbonate and 67% using nahcolite was obtained (bed depth, 36 in.; temperature, 300°F). NO_x removal of 42% has been reported for the process. If bicarbonate is not regenerated, the full flyash loading of the gas stream can be collected in a baghouse along with the reaction product, Na₂SO₄. Nahcolite deposits do not occur widely.

6 COMMERCIAL FGD IN THE U.S.

6.1 OPERATIONAL PROCESSES^{42,49,50,61,73,76}

6.1.1 Limestone Slurry - Throwaway Process

A generalized limestone scrubbing process has already been described. Fig. 6.1 shows a more detailed flow diagram of this process. Stack gas is contacted with an 8-17% solids slurry of limestone. The basic reaction between the calcium and sulfite ions in the absorber produces insoluble calcium sulfite hemihydrate.

Some oxidation of sulfite to sulfate occurs and gypsum is formed.

Currently, eight commercial-sized limestone units are in operation in the U.S. At the Sherburne plant, the limestone is supplemental to the boiler's alkaline flyash.

Unit	FGD Size (MW)	Utility	Scrubber Vendor
Lawrence 4	125	Kansas Power and Light	Combustion Engineering
Lawrence 5	400	Kansas Power and Light	Combustion Engineering
Will County 1	167	Commonwealth Edison	Babcock and Wilcox
La Cygne 1	820	Kansas City Power and Light	Babcock and Wilcox
Cholla 1	115	Arizona Public Service	Research-Cottrell
Sherburne 1	700	Northern States Power	Combustion Engineering
Sherburne 2	700	Northern States Power	Combustion Engineering
Widows Creek 8	550	Tennessee Valley Authority	Tennessee Valley Authority

FIG. 6.1. LIMESTONE SLURRY PROCESS FLOW DIAGRAM

6.1.2 Lime Slurry - Throwaway Process

In the lime slurry process (Chemico process shown in Fig. 6.2), the reaction of slaking is added to the scrubber reactions described in the limestone process.

The pH of lime slurries is higher than that of limestone slurries because of the additional hydroxide ion from the slaked lime. Due to the high pH, precipitation of CaCO_3 is possible.

One commercial operation, Colstrip units 1 and 2 of Montana Power Company, and one demonstration system, Four Corners unit 5A of Arizona Public Service, use alkaline flyash for SO_2 removal with the addition of lime to maintain alkalinity.

At present, 10 lime units are commercially operational in the U.S. They include carbide lime at the Cane Run Station and lime-alkaline flyash at the Colstrip Station.

Unit	FGD Size (MW)	Utility	Scrubber Vendor
Hawthorn 3	90	Kansas City Power and Light	Combustion Engineering
Hawthorn 4	90	Kansas City Power and Light	Combustion Engineering
Cane Run 4	178	Louisville Gas and Electric	American Air Filter
Phillips 1-6	390	Duquesne Light	Chemico
Elrama 1-2	200	Duquesne Light	Chemico
Green River 1-3	64	Kentucky Utilities	American Air Filter
Colstrip 1	360	Montana Power	Combustion Equipment
Colstrip 2	360	Montana Power	Combustion Equipment
Bruce Mansfield 1	835	Pennsylvania Power	Chemico
Conesville 5	400	Columbus and Southern	UOP

FIG. 6.2. LIME SLURRY PROCESS FLOW DIAGRAM

6.1.3 Na Solution - Throwaway Process

Fig. 6.3 outlines a throwaway process using an Na scrubbing solution to remove SO₂. This type is in operation at the Reid Gardner Station of Nevada Power Company, Combustion Equipment Associates, vendor.

Particulate removal is achieved by multicyclones ahead of a venturi scrubber. Scrubber bypass is provided downstream of the multicyclones. Powdered Na₂CO₃ is dissolved in a mixing tank in recycled water from the ash pond. Insolubles are settled out in a clarifier and the Na₂CO₃ solution is pumped to the venturi where it contacts the hot flue gas. Some SO₂ removal occurs in the venturi along with humidification and cooling of the gas. From the venturi, the partially scrubbed gas enters a single-tray absorber tower. Flowing upward, the gas bubbles through the single sieve tray which is flooded with clarified ash pond water. Cleaned gas exits the tower through a mist eliminator. Air which has been indirectly heated with steam is mixed with the gas from the scrubber to provide plume buoyancy from the stack.

Clarifier underflow and a bleed stream of scrubber effluent are mixed with alkali in a neutralizing tank. The spent liquor, insolubles, and particulates from this tank are pumped to a settling pond whose overflow goes to an evaporation pond. There is no recycle to the system from these ponds.

The reaction of Na₂CO₃ and SO₂ to form Na₂SO₃ is primary:

A small amount of 2NaHSO₃ also forms:

6.2 COMMERCIAL FGD INSTALLATIONS

6.2.1 Process Capacities and Plant Installations^{48-50,61,73,76,85,92}

An overview of FGD capacity in the U.S. is given in Table 6.1. The 21 commercial units in operation have a total capacity of 6,900 MW; the 28 units under construction have 11,000 MW; and the 19 planned units have 9,300 MW. Total current commercial involvement is 27,200 MW. Table 6.1 includes a similar portrayal of FGD capacity for the plant development units which were described in Sec. 5. Their total capacity is 1100 MW.

FIG. 6.3. SODIUM SCRUBBING THROWAWAY PROCESS FLOW DIAGRAM

Table 6.1. Capacity by Process of FGD Installations in the U.S. - May 31, 1977

Process	Commercial Units								Demonstration and Prototype Units								Total Capacity	
	In Operation		Under Constr		Planned		Total		Idle		In Operation		Constr Plan		Total		No.	MW
	No.	MW	No.	MW	No.	MW	No.	MW	No.	MW	No.	MW	No.	MW	No.	MW		
I. Throwaway																		
A. Nonregenerable																		
Limestone	6	2,177	17	5,714	8	3,985	31	11,876	1	163	1	50	-	-	2	213	33	12,089
Lime	8	2,247	7	3,533	3	1,635	18	7,415	-	-	2	225	-	-	2	225	20	7,640
Lime/limestone	-	-	-	-	-	-	-	-	-	-	2	20	-	-	2	20	2	20
Lime/alkaline flyash	2	720	1	450	4	2,490	7	3,660	-	-	-	-	-	-	-	-	7	3,660
Limestone/alkaline ash	2	1,400	-	-	-	-	2	1,400	-	-	-	-	-	-	-	-	2	1,400
Sodium carbonate	3	375	-	-	2	575	5	1,050	-	-	-	-	-	-	-	-	5	1,050
B. Regenerable																		
Double alkali	-	-	1	575	1	250	2	825	1	20	-	-	1	277	2	297	4	1,122
Chiyoda 101	-	-	-	-	-	-	-	-	1	23	-	-	-	-	1	23	1	23
II. Recoverable																		
A. Regenerable																		
Wellman-Lord	-	-	2	715	-	-	2	715	-	-	1	115	-	-	1	115	3	830
Magnesium oxide	-	-	-	-	1	240	1	240	-	-	1	120	-	-	1	120	2	360
Activated carbon	-	-	-	-	-	-	-	-	1	20	-	-	-	-	1	20	1	20
Molten carbonate	-	-	-	-	-	-	-	-	1	20	-	-	-	-	1	20	1	20
Aqueous carbonate	-	-	-	-	-	-	-	-	-	-	-	-	1	- ^a	1	- ^a	1	- ^a
Sodium citrate	-	-	-	-	-	-	-	-	-	-	-	-	1	60	1	60	1	60
Total	21	6,919	28	10,987	19	9,275	68	27,181	5	246	7	530	3	337	15	1,113	83	28,294

^aIndefinite.

All units in commercial operation have a throwaway product and a non-regenerable sorbent, but one double-alkali unit is under construction and one is planned. The W-L process (recoverable product and regenerable sorbent) is being installed at two commercial units under construction and a MgO slurry process (also recoverable, regenerable) is being planned for one unit.

The specifications and features of the commercial FGD installations of Table 6.1 are detailed in Table 6.2 (in operation), Table 6.3 (under construction), and Table 6.4 (planned). Installations are listed chronologically by utility, and by station within the utility. Features, such as fuel characteristics, are given on a nominal basis because full information would involve multiple entries.

The equipment arrangements for FGD installations in operation (Table 6.2) are illustrated in Figs. 6.4-6.12. These diagrams include the major types of particulate scrubbers and of SO₂ absorbers in current use.

6.2.2 Operational Plant Experience^{48-50,56-61,73-78,85-92}

Lawrence 4-5

The Kansas Power and Light Company's Lawrence installation was the initial system cited by EPA as evidence of demonstrated technology. A Combustion Engineering limestone injection - wet scrubber system was retrofitted to the 125-MW unit 4 and went into service in the fall of 1968. The same system was included as original equipment when the 400-MW unit 5 was designed; operation of this system began in 1971.

Limestone and coal were pulverized together and were injected into the boiler of unit 4 as a mixture. A wet scrubber (marble bed) was placed downstream of the air heater for simultaneous removal of flyash and SO₂. Particulate removal was good but heavy scrubber scaling required a reduction in limestone addition and SO₂ removal decreased to 55-60%. Modification in limestone injection to the boiler and improved cleaning techniques were used to overcome the problems of plugging, scaling, and corrosion. Further modifications included the addition of a high-solid slurry crystallization process to both units 4 and 5. This reduced cleaning requirements and increased scrubber availability. In the fall of 1974, the furnaces began burning low-sulfur coal

Table 6.2. Commercial FGD Systems Installed in the U.S.

Item No.	Power Plant			Total MW	Startup Date	Vendor ^b	FGD Installation				Total Trains	Fuel			
	Utility	Station	Boiler No.				Process	New Retrofit	Throw-away	Bypass		Type	Btu/lb	% S	% Ash
1	Kansas Power and Light	Lawrence	4	125	(12/68) 1/77	CE	Limestone scrubbing	R	Yes	Yes	1	Coal (gas-oil)	10,000	0.5	11
2	Kansas Power and Light	Lawrence	5	400	(11/71) 6/78	CE	Limestone to boiler	N	Yes	Yes	8, 2	Coal (gas-oil)	10,000	0.5	11
3	Commonwealth Edison	Will County	1	167	2/72	B & W	Limestone scrubbing	R	Yes	Yes	2	Coal	9,463	2.0-3.5	15
4	Kansas City Power and Light	Hawthorn	4	90	(8/72) 1/77	CE	Lime scrubbing	R	Yes	Yes	2	Coal (gas)	9,800-11,400	0.6-3.0	11-14
5	Kansas City Power and Light	Hawthorn	3	90	(11/72) 1/77	CE	Lime scrubbing	R	Yes	Yes	2	Coal (gas)	9,800-11,400	0.6-3.0	11-14
6	Kansas City Power and Light	La Cygne	1	820	2/73	B & W	Limestone scrubbing	N	Yes	No	8	Coal	9,000-9,700	3.0-6.0	15-25
7	Duquesne Light	Phillips	1-6	390	7/73	Chemico	Lime scrubbing	R	Yes	Yes	4	Coal	11,350	1.0-2.8	18.2
8	Duquesne Light	Elrama	1-2	200	10/75, 2/76	Chemico	Lime scrubbing	R	Yes	Yes	5	Coal	11,350	1.0-2.8	18.2
9	Arizona Public Service	Cholla	1	115	10/73	R-C	Limestone scrubbing	R	Yes	Yes	1	Coal	12,146	0.4-1.0	12.3
10	Nevada Power	Reid Gardner	2	125	12/73	CEA	Sodium carbonate scrubbing	R	Yes	Yes	1	Coal	12,450	0.5-1.0	8-10
11	Nevada Power	Reid Gardner	1	125	4/74	CEA	Sodium carbonate scrubbing	R	Yes	Yes	1	Coal	12,450	0.5-1.0	8-10
12	Nevada Power	Reid Gardner	3	125	7/76	CEA	Sodium carbonate scrubbing	N	Yes	Yes	1	Coal	12,450	0.5-1.0	8-10
13	Kentucky Utilities	Green River	1-3	64	9/75	AAF	Lime scrubbing	R	Yes	Yes	1	Coal	11,000	3.5-4.0	14
14	Montana Power	Colstrip	1	360	10/75	CEA/ADL	Lime/alkaline flyash scrubbing	N	Yes	No	3	Coal	8,800	0.8	-
15	Montana Power	Colstrip	2	360	7/76	CEA/ADL	Lime/alkaline flyash scrubbing	N	Yes	No	3	Coal	8,800	0.8	-
16	Northern States Power	Sherburne	1	700	3/76	CE	Limestone/alkaline flyash scrubbing	N	Yes	No	12	Coal	8,520	0.8	9
17	Northern States Power	Sherburne	2	700	4/77	CE	Limestone/alkaline flyash scrubbing	N	Yes	No	12	Coal	8,300	0.8	9
18	Pennsylvania Power	Bruce Mansfield	1	835	4/76	Chemico	Lime scrubbing	N	Yes	No	6	Coal	11,900	3.0-4.5	8-15
19	Louisville Gas and Electric	Cane Run	4	178	8/76	AAF	Carbide lime scrubbing	R	Yes	Yes	2	Coal	11,500	3.5-4.0	11-12
20	Columbus & So. Ohio Electric	Conesville	5	400	2/77	UOP	Lime scrubbing	N	Yes	Yes	2	Coal	10,800	4.5-4.9	15
21	Tennessee Valley Authority	Widows Creek	8	550	5/77	TVA	Limestone scrubbing	R	Yes	Yes	4	Coal	10,000	3.7	25

(continued)

Table 6.2 (continued)

Item No.	SIP ^c	Pollutant Control						FGD Product		Flue Gas Reheat		
		Particulate			SO ₂			Material	Disposal	Mode	Heat Source	Temp. to Stack, °F
		Device	Make ^b	%	Device	Make ^a	%					
1	1.50 ²	ESP, venturi rod	CE	99	Spray tower	CE	75	Unstabilized sludge	Unlined pond	Inline	Steam	150-155
2	1.50 ²	Single-stage marble bed	CE	99	Single-stage marble bed	CE	65	Unstabilized sludge	Unlined pond	Inline	Steam	150-155
3	1.80 ¹	ESP, variable-throat venturi	JW, B & W	98	Tray absorber	B & W	82	Stabilized sludge	Clay-lined pond	Inline	Steam	175
4	0.7 ⁹	Marble bed	CE	99	Marble bed	CE	70	Unstabilized sludge	Unlined pond	Inline	Steam	180
5	0.7 ⁹	Marble bed	CE	99	Marble bed	CE	70	Unstabilized sludge	Unlined pond	Inline	Steam	180
6	1.50 ⁴	Variable-throat venturi	B & W	98	Perforated plate	B & W	72	Unstabilized sludge	Unlined pond	Inline	Air/steam	175
7	0.7 ⁷	ESP, variable-throat venturi	Chemico	99	A single-stage and 1 dual-stage venturi	Chemico	83	Stabilized sludge	Unlined pond	Direct	Oil	125
8	0.67 ²	ESP, venturi scrubber	Chemico	-	Single-stage venturi scrubber	Chemico	83	Stabilized sludge	Unlined pond	-	-	120
9	1.00 ¹	Flooded disc venturi scrubber	R-C	99.7	Packed tower	R-C	78	Unstabilized sludge	Unlined pond	Inline	Steam	181
10	0.15 ⁴	Multicyclone and 2 parallel venturi scrubbers	CEA	99.4	Single-stage wash tray	CEA	85	Spent liquor	Unlined pond	Direct	Hot air	165
11	0.15 ⁴	Multicyclone and 2 parallel venturi scrubbers	CEA	-	Single-stage wash tray	CEA	85	Spent liquor	Unlined pond	Direct	Hot air	165
12	0.15 ⁴	Multicyclone and 2 parallel venturi scrubbers	R-C, CEA	79	Single-stage wash tray	CEA	85	Spent liquor	Unlined pond	Direct	Hot air	165
13	2-5 ¹	Variable-throat venturi scrubber	AAF	99.7	Ten-compartment mobile bed	AAF	80	Stabilized sludge	Unlined pond	None	None	-
14	2.00 ⁴	Variable-throat venturi scrubber	CEA	99	Variable-throat venturi scrubber	CEA	88	Unstabilized sludge	Unlined pond	Inline	Steam	180
15	2.00 ⁴	Variable-throat venturi scrubber	CEA	99	Variable-throat venturi scrubber	CEA	88	Unstabilized sludge	Unlined pond	Inline	Steam	180
16	2.00 ⁶	Fixed and adjustable rod venturi	CE	99	Single-stage marble bed	CE	50	Oxidized, unstabilized sludge	Lined pond	Inline	Water	170
17	0.60 ²	Fixed rod venturi	CE	99	Single-stage marble bed	CE	50	Unstabilized sludge	Lined pond	Inline	Water	170
18	0.60 ²	Variable-throat venturi scrubber	Chemico	99.8	Fixed-throat venturi scrubber	Chemico	92	Stabilized sludge	Unlined pond	Direct	Oil	160 (design)
19	1.20 ¹	ESP	-	99	Mobile bed contractor	AAF	90	Stabilized sludge	Unlined pond	None	None	-
20	-	ESP	UOP	99.5	Two-bed TCA	UOP	89.6	Stabilized sludge	Lined pond	None	None	-
21	4.00 ²	ESP, variable-throat venturi scrubber	-	99	Grid tower or TCA	UOP	75	Unstabilized sludge	Pond	Air injection	Steam	175

^aDates in parentheses refer to first installation.^bSee Appendix B.^cSee Appendix C.

Table 6.3. Commercial FGD Installations under Construction in the U.S.

Item No.	Power Plant			Boiler No.	Total MW	Startup Date	FGD Installation					Total Trains	Fuel			
	Utility	Station	Vendor				Process	New/Retrofit	Throw-away	Bypass	Type		Btu/lb	% S	% Ash	
1	South Carolina Public Service	Winyah	2	140	6/77	B & W	Limestone scrubbing	N	Yes	Yes	1	Coal	11,800	1.1	12-13	
2	Arizona Public Service	Cholla	2	250	6/77	R-C	Limestone scrubbing	N	Yes	Yes	4	Coal	10,000	0.44-1.0	12	
3	Utah Power and Light	Huntington	1	415	6/77	Chemico	Lime scrubbing	N	Yes	-	-	Coal	-	0.5	-	
4	Louisville Gas and Electric	Mill Creek	3	425	7/77	AAF	Lime scrubbing	N	Yes	Yes	-	Coal	11,000	3.5-4.0	-	
5	Louisville Gas and Electric	Mill Creek	4	425	7/79	AAF	Lime scrubbing	N	Yes	-	-	Coal	-	3.5-4.0	-	
6	Louisville Gas and Electric	Cane Run	5	183	12/77	CE	Lime scrubbing	R	Yes	Yes	-	Coal	13,000	3.5-4.0	-	
7	Springfield City Utilities	Southwest	1	200	8/77	UOP	Limestone scrubbing	N	Yes	Yes	2	Coal	12,500	3.5	12-14	
8	Minnesota Power Cooperative	Milton R. Young	2	450	8/77	CEA/ADL	Lime/alkaline flyash scrubbing	N	Yes	-	2	Lignite	6,500	0.7	-	
9	Indianapolis Power and Light	Petersburg	3	530	9/77	UOP	Limestone scrubbing	N	Yes	-	4	Coal	-	3.0-3.5	-	
10	Pennsylvania Power	Bruce Mansfield	2	835	10/77	Chemico	Lime scrubbing	N	Yes	No	6	Coal	11,900	3.0-4.5	8-15	
11	Texas Utilities	Martin Lake	1	750	10/77	R-C	Limestone scrubbing	N	Yes	Yes	6	Coal	7,380	1.0	-	
12	Texas Utilities	Martin Lake	2	750	3/78	R-C	Limestone scrubbing	N	Yes	Yes	6	Coal	7,380	1.0	-	
13	Texas Utilities	Monticello	3	750	3/78	Chemico	Limestone scrubbing	N	Yes	Yes	3	Lignite	-	1.0	-	
14	Public Service Company of New Mexico	San Juan	1	375	11/77	DP/AC	Na solution/SO2 reduction	N	No	-	4	Coal	-	0.8	-	
15	Public Service Company of New Mexico	San Juan	2	340	11/77	DP/AC	Na solution/SO2 reduction	R	No	-	4	Coal	-	0.8	18	
16	So. Mississippi Electric Power	R. D. Morrow	1	180	11/77	EV/RS	Limestone scrubbing	N	Yes	-	-	Coal	-	1.0	-	
17	So. Mississippi Electric Power	R. D. Morrow	2	180	6/78	EV/RS	Limestone scrubbing	N	Yes	-	-	Coal	-	1.0	-	
18	Southern Illinois Power Cooperative	Marion No. 4	4	184	1/78	B & W	Limestone scrubbing	N	Yes	-	2	Coal	-	-	-	
19	Alabama Electric Cooperative	Tombigbee	2	160	3/78	PB	Limestone scrubbing	N	Yes	-	2	Coal	-	0.8-1.5	-	
20	Alabama Electric Cooperative	Tombigbee	3	160	3/79	PB	Limestone scrubbing	N	Yes	-	2	Coal	-	0.8-1.5	-	
21	Arizona Electric Power Cooperative	Apache	2	200	6/78	R-C	Limestone scrubbing	N	Yes	-	2	Coal	10,200	0.5-0.8	-	
22	Arizona Electric Power Cooperative	Apache	3	200	6/79	R-C	Limestone scrubbing	N	Yes	-	2	Coal	10,200	0.5-0.8	-	
23	Kansas Power and Light	Jeffery	1	340	6/78	CE	Limestone scrubbing	N	Yes	Yes	-	Coal	8,000	0.32	6	
24	Kansas Power and Light	Jeffery	2	340	6/79	CE	Limestone scrubbing	N	Yes	Yes	-	Coal	8,000	0.32	6	
25	Central Illinois Public Service	Newton	1	575	7/78	BU/EN	Sodium-lime double alkali	N	Yes	Yes	4	Coal	10,900	4.0-0.2% Cl	-	
26	Central Illinois Light	Duck Creek	1	400	8/78	RS/EV	Limestone scrubbing	N	Yes	No	4	Coal	-	0.6	-	
27	Allegheny Power System	Pleasants	1	625	3/79	B & W	Lime scrubbing	N	Yes	-	4	Coal	-	4.5 max	-	
28	Allegheny Power System	Pleasants	2	625	3/80	B & W	Lime scrubbing	N	Yes	-	4	Coal	-	4.5 max	-	

(continued)

Table 6.3 (continued)

Item No.	SIP ^b	Pollutant Control						FGD Product		Flue Gas Reheat		
		Particulate			SO ₂			Material	Disposal	Mode	Heat Source	Temp. to Stack, °F
		Device	Make ^a	%	Device	Make ^a	%					
1	-	ESP	R-C	99.4	-	B & W	69	Unstabilized sludge	Unlined pond	None	None	-
2	1.00 ¹	Venturi scrubber	R-C	70	Packed tower	R-C	-	-	-	-	-	-
3	1.00 ⁶	ESP	-	-	-	Chemico	-	Stabilized sludge	-	-	-	-
4	1.20 ¹	-	-	-	Mobile-bed contactor	AAF	-	Stabilized sludge	-	-	-	-
5	1.20 ¹	-	-	-	-	-	-	-	-	-	-	-
6	1.10 ¹	-	-	-	Mobile bed	-	-	-	-	-	-	-
7	-	ESP	-	-	Three-bed TCA	UOP	-	Stabilized sludge	-	Direct	Steam	-
8	3.00 ²	ESP	-	99.6	-	-	75	-	-	-	-	-
9	1.20 ¹	ESP	R-C	99.3	TCA	-	85	Stabilized sludge	-	-	-	-
10	0.60 ²	Variable-throat venturi scrubber	Chemico	99.8	Fixed-throat venturi scrubber	Chemico	92	Stabilized sludge	Unlined pond	Direct oil	-	-
11	0.20 ⁹	ESP	-	-	-	R-C	-	Stabilized sludge	-	Direct	Bypassed flue gas	-
12	0.20 ⁹	ESP	-	-	-	R-C	-	Stabilized sludge	-	Direct	Bypassed flue gas	-
13	0.20 ⁹	ESP	-	-	-	-	-	-	-	-	-	-
14	3M ¹⁰	ESP, venturi scrubber	-	99.7	-	-	-	Sulfur	Market	-	-	-
15	3M ¹⁰	ESP, venturi scrubber	-	99.7	-	-	-	Sulfur	Market	-	-	-
16	-	ESP	EV	99.6	-	RS	85	Stabilized sludge	-	-	-	-
17	-	-	-	99.6	-	RS	85	Stabilized sludge	-	-	-	-
18	-	-	-	-	-	-	89.4	-	-	-	-	-
19	4.00 ²	ESP	-	-	-	-	-	-	-	-	-	-
20	4.00 ²	ESP	-	-	-	-	-	-	-	-	-	-
21	1.00 ¹	ESP	UOP	99.4	Packed tower	-	85	-	-	-	-	-
22	1.00 ¹	ESP	UOP	99.4	Packed tower	-	85	-	-	-	-	-
23	1.50 ⁴	ESP	CE	90	Spray tower	CE	60	-	-	Direct	Bypassed flue gas	-
24	1.50 ⁴	ESP	-	-	Spray tower	CE	-	-	-	-	-	-
25	6.00 ¹	ESP, cocurrent spray precoolers	BU	99.5	Two-tray mobile ball scrubber	-	95	Stable filter cake	-	-	-	-
26	1.20 ¹	ESP	-	99.8	Packed-bed venturi	RS	75	-	-	-	-	-
27	-	ESP, venturi scrubber	-	99.5	Perforated plate reaction tray	-	90	Stabilized sludge	-	-	-	-
28	-	ESP, venturi scrubber	-	99.5	Perforated plate reaction tray	-	90	Stabilized sludge	-	-	-	-

^a See Appendix B.^b See Appendix C.

Table 6.4. Commercial FGD Installations Planned in the U.S.

Item No.	Power Plant			Boiler No.	Total MW	Startup Date	Vendor ^a	FGD Installation				Total Trains	Fuel			
	Utility	Station						Process	New/ Retrofit	Throw-away	Bypass		Type	Btu/lb	% S	% Ash
1	Columbus & South Ohio Electric	Conesville	6	400	-	UOP	Lime scrubbing	N	Yes	Yes	2	Coal	10,800	4.5-4.9	15	
2	Utah Power and Light	Emery	1	400	-	Chemico	Lime scrubbing	N	Yes	-	-	Coal	-	0.5	-	
3	Cooperative Power Association	Coal Creek	1	545	-	CE	Lime/alkaline flyash	N	Yes	Yes	4	Lignite	6,300	0.63	7	
4	Cooperative Power Association	Coal Creek	2	545	-	CE	Lime/alkaline flyash	N	Yes	Yes	4	Lignite	6,300	0.63	7	
5	Texas Utilities	Martin Lake	3	750	-	R-C	Limestone scrubbing	N	Yes	Yes	6	Coal	7,380	1.0	-	
6	Texas Utilities	Martin Lake	4	750	-	R-C	Limestone scrubbing	N	Yes	Yes	6	Coal	7,380	1.0	-	
7	Pennsylvania Power	Bruce Mansfield	3	835	-	PK	Lime scrubbing	N	Yes	No	5	Coal	11,900	3.0-4.5	-	
8	Southern Indiana Gas and Electric	A. B. Brown	1	250	-	FMC	Sodium-lime double alkali	N	Yes	-	-	Coal	-	3.75	-	
9	Salt River Project	Coronado	1	280	-	PK	Limestone scrubbing	N	Yes	Yes	2	Coal	-	1.0 max	-	
10	Salt River Project	Coronado	2	280	-	PK	Limestone scrubbing	N	Yes	Yes	2	Coal	-	1.0 max	-	
11	Pacific Power and Light	Jim Bridger	4	550	-	UOP	Sodium carbonate scrubbing	N	Yes	Yes	3	Coal	9,000	0.56	-	
12	Commonwealth Edison	Powerton	5	425	-	UOP	Limestone scrubbing	R	Yes	-	-	Coal	10,500	3.6	8.3	
13	Brazos Electric Power Cooperative	Atacosa McMullen	1	400	-	B & W	Limestone scrubbing	N	Yes	-	-	Lignite	-	1.67	-	
14	Basin Electric Power Cooperative	Missouri Basin	1	550	-	R-C	Limestone scrubbing	N	Yes	-	-	Lignite	8,000	0.8	7	
15	Basin Electric Power Cooperative	Missouri Basin	2	550	-	R-C	Limestone scrubbing	N	Yes	-	-	Lignite	8,000	0.8	7	
16	Philadelphia Electric	Eddystone	1B	240	-	UE/PECO	Magnesium oxide scrubbing	R	No	-	-	Coal	-	2.5	-	
17	Montana Power	Colstrip	3	700	-	CEA/ADL	Lime/alkaline flyash scrubbing	N	Yes	No	3	Coal	-	0.7	-	
18	Montana Power	Colstrip	4	700	-	CEA/ADL	Lime/alkaline flyash scrubbing	N	Yes	No	3	Coal	-	0.7	-	
19	Nevada Power	Reid Gardner	4	125	-	CEA	Sodium carbonate scrubbing	N	Yes	-	-	Coal	-	0.5	-	

(continued)

Table 6.4 (continued)

Item No.	SIP ^b	Pollutant Control						FGD Product		Flue Gas Reheat		
		Particulate			SO ₂			Material	Disposal	Mode	Heat Source	Temp. to Stack, °F
		Device	Make ^a	%	Device	Make ^a	%					
1	-	ESP	UOP	99.5	Two-bed TCA	UOP	89.6	Stabilized sludge	Lined pond	None	None	-
2	1.00 ⁶	ESP	EN	-	-	-	80	Stabilized sludge	-	-	-	-
3	1.20 ¹	ESP	-	99.5	Spray tower	CE	90	Unstabilized sludge	Pond	Direct	Bypassed flue gas	210
4	1.20 ¹	ESP	-	99.5	Spray tower	CE	90	Unstabilized sludge	Pond	Direct	Bypassed flue gas	210
5	0.2 ⁹	-	-	-	-	-	-	-	-	Direct	Bypassed flue gas	-
6	0.2 ⁹	-	-	-	-	-	-	-	-	Direct	Bypassed flue gas	-
7	0.60 ²	ESP	-	-	Weir horizontal spray scrubber	PK	-	Stabilized sludge	Unlined pond	Direct	Fuel oil	160
8	-	ESP	-	-	Dian contactor	-	-	Stable filter cake	Landfill	-	-	-
9	-	ESP	JW	-	Weir horizontal spray scrubber	PK	-	-	-	Direct	Bypassed flue gas	-
10	-	ESP	JW	-	Weir horizontal spray scrubber	PK	-	-	-	Direct	Bypassed flue gas	-
11	0.1 ⁹	ESP	-	-	Sieve-tray scrubber	UOP	91	Spent liquor	Lined pond	None	None	-
12	1.00 ¹	-	-	-	TCA	UOP	-	Stabilized sludge	-	-	-	-
13	-	ESP	-	-	-	-	86	-	-	-	-	-
14	-	ESP	-	-	-	-	-	-	-	-	-	-
15	-	ESP	-	-	-	-	-	-	-	-	-	-
16	0.60 ²	ESP, venturi scrubber	-	-	-	-	-	Sulfuric acid	Market	-	-	-
17	1.00 ⁴	Variable-throat venturi scrubber	CEA	99	Variable-throat venturi scrubber	CEA	88	Unstabilized sludge	Unlined pond	Inline	Steam	180
18	1.00 ⁴	Variable-throat venturi scrubber	CEA	99	Variable-throat venturi scrubber	CEA	88	Unstabilized sludge	Unlined pond	Inline	Steam	180
19	0.15 ⁴	-	-	-	-	-	-	-	-	-	-	-

^aSee Appendix B.^bSee Appendix C.

FIG. 6.4. FGD EQUIPMENT ARRANGEMENT AT LAWRENCE 4 & 5 (1968)
AND HAWTHORN 3 & 4 STATIONS

FIG. 6.5. FGD EQUIPMENT ARRANGEMENT AT LAWRENCE 4 & 5 (1977)
AND WILL COUNTY 1 STATIONS

FIG. 6.6. FGD EQUIPMENT ARRANGEMENT AT LA CYGNE I
AND SHERBURNE COUNTY I STATIONS

FIG. 6.7. FGD EQUIPMENT ARRANGEMENT AT PHILLIPS 1-6 AND
BRUCE MANSFIELD 1 STATIONS

FIG. 6.8. FGD EQUIPMENT ARRANGEMENT AT ELRAMA 1-2 STATION

FIG. 6.9. FGD EQUIPMENT ARRANGEMENT AT CHOLLA I AND WIDOWS CREEK 8 STATIONS

FIG. 6.10. FGD EQUIPMENT ARRANGEMENT AT GREEN RIVER 1-3
AND REID GARDNER 1, 2 & 3 STATIONS

FIG. 6.11. FGD EQUIPMENT ARRANGEMENT AT COLSTRIP 1,2

FIG. 6.12. FGD EQUIPMENT ARRANGEMENT AT CANE RUN 4 AND CONESVILLE 5 STATIONS

from Wyoming (10,000 Btu/lb, 0.5% sulfur, and 11% ash). Scrubber system operation improved somewhat and became more economical with the decrease in sulfur to be removed.

The unit 4 scrubbing system has recently been replaced with limestone scrubbing in a spray tower. The new system went on stream in January 1977. Unit 5 is presently undergoing the same type scrubber changes and should be in operation in mid-1978.

The plant has no sludge treatment or fixation systems. Sludge from the FGD units is pumped to unlined ponds for settling. The pond for water clarification receives cooling tower blowdown and makeup water from the Kansas River as well as the scrubber liquor. Clarified water from this pond is recycled.

Will County

An existing 167-MW unit was equipped with a two-train Babcock and Wilcox Company (B & W) limestone scrubbing system and was started up in the spring of 1972. Originally, one scrubber was a two-stage sieve tray module while the second was a Turbulent Contact Absorber (TCA). After a year's operation, the TCA was shut down and efforts were concentrated on improving the performance of the other train. Scaling, plugging, and corrosion prevented sustained operation of the single train until 1974. Use of fresh water for washing and discharge of wash water from the scrubbing system have since reduced the scaling and plugging problems. Operating problems were reduced through the burning of low-sulfur coal. The second train's scrubber was modified to a two-stage sieve tray module and was returned to service in the spring of 1975. However, the scrubber problems returned when high-sulfur coal was burned during a test run. Mist eliminators in both scrubbers became almost completely plugged and heavy scaling occurred in the second scrubber. Commonwealth Edison will abandon the SO₂ removal effort at Will County and in the future use the scrubbers for particulate removal only, burning low-sulfur coal in the boiler.

Sludge, stabilized with lime and flyash, is transported in concrete mixing trucks to a dumping site. Some treated sludge has been used as fill for a future golf course.

Hawthorn 3-4

Hawthorn 4 originally involved furnace injection of limestone by a Combustion Engineering system which was installed in 1972. Because of boiler tube fouling, the system was converted in early 1973 for injection of limestone just ahead of the scrubber. Plugging, scaling, and erosion of the system resulted in low availability. A decision to burn low-sulfur coal was made in an effort to improve dependability. Modifications to the system continue.

Hawthorn 3 was operated on limestone furnace injection followed by wet scrubbing. Problems similar to those encountered at unit 4 led to a recommendation to change to a postboiler injection system or to lime injection.

In 1976 the utility began the conversion of both units 3 and 4 from injection and tail-end scrubbing to a lime slurry scrubbing system. Startup of the scrubbing system on unit 4 was in January and that on unit 3 was in February 1977.

La Cygne 1

La Cygne unit 1 has a new (1973) 820-MW boiler designed for burning locally mined coal with high ash (15-25%) and high sulfur (5.3-6%). The limestone-slurry FGD system has seven identical venturi-sieve tray (perforated-plate) modules for particulate and SO₂ removal, and an eighth module will be added during 1977. Its addition is intended to increase cruising capability of the unit from 700 to 800 MW. There is no flue gas bypass of the system but each module can be isolated with dampers for maintenance needs.

Various combinations of scaling, plugging, and corrosion have created major problems throughout the FGD system, requiring an operating-maintenance force which has gradually decreased from 100 to 51 people. However, a strenuous development program has reduced some problems. Scaling on the perforated plates has been lessened by careful control of the slurry pH at 5.55. Demister plugging has been reduced by changing the demister washing arrangement to a continuous, topside wash which is countercurrent to the gas flow. Reheaters and fans are washed frequently and the amount of reheating has been increased to at least 50°F. Corrosion-resistant metals and coatings are

being tested and used where promising, but other problems persist. For example, the process differences between the gray and brown limestone from the same local quarry are unanswered.

No sludge fixation is used or intended. Sludge and flyash are deposited in an unlined pond and most of the clarified water is recycled to the open-loop system.

Phillips 1-6, Elrama 1-2

The Phillips Power Station, Duquesne Light Company, consists of six coal-fired boilers with a combined rating of 390 MW. Four trains of venturi scrubbers using lime as reagent were retrofitted to the boiler units. Three trains are single-stage scrubbers originally designed for particulate removal; the fourth train is a two-stage venturi installed as a test unit for determining design feasibility for SO₂ emission compliance. Operation of a part of the scrubbing system began in July 1973.

Early problems -- equipment erosion and corrosion, ID fan stress, acid condensate leakage in the stack -- caused scrubber system outages. After an extended outage, scrubbers were restarted in March 1974. Subsequent problems included scrubber scale and other deposits, ID fan deposits, high water usage in open loop, and sludge disposal. The Phillips scrubber system has been in operation since March 1974 with various combinations of boilers and scrubbers.

The Elrama Station consists of four coal-fired boilers with a total rating of 500 MW. Five single-stage scrubber trains using lime were retrofitted to the four boilers but only two boilers have been connected to the FGD system. One boiler was connected to it in October 1975 and a second was connected in February 1976. There have been problems with recycle pump erosion/corrosion and with scrubber scale and deposits.

Sludge fixation and disposal tests were conducted at Phillips by Dravo, IU Conversion Systems, Inc. (IUCS), and U.S. Utilities Service, and at Elrama by IUCS. The IUCS system will be used at both stations.

By December 1977, construction should be completed on the total systems at both locations for SO₂ scrubbing with MgO-modified quicklime.

Cholla 1

A two-train limestone scrubbing system designed by Research-Cottrell was retrofitted to a 115-MW boiler at Cholla Station, Arizona Public Service Company. Each train consists of a flooded-disc scrubber, packed-tower absorber (only one absorber actually contains packing), mist eliminator, and reheater. Operation of the system began in December 1973 with low-sulfur coal.

Corrosion was experienced in the reheater due to condensation in the lead-in ductwork. Other problems have included wearing of control valves, scaling on the packing, and plugging of nozzles and demisters. High on-stream time has now been achieved, but extensive maintenance and operation efforts are required.

The plant has no sludge treatment or fixation systems. Scrubber sludge is pumped alternately with the flyash through common pipes to flyash ponds which have high evaporation and high ground absorption. No water is recycled to the FGD system from the present ponds but lined ponds now under construction will provide some return water.

A second unit, 250 MW, is now under construction with Research-Cottrell's limestone scrubbing system being installed for desulfurization.

Reid Gardner 1-3

Three identical 125-MW boilers at Reid Gardner Station, Nevada Power Company, are equipped with Na_2CO_3 scrubbing systems installed by Combustion Equipment Associates. Two units are retrofit; unit 3 is a new installation. The system, including a venturi scrubber followed by a tray absorber, was designed to use naturally occurring trona as the sorbent, but market conditions have favored the use of soda ash. Startup of unit 1 was in March 1974.

The multicyclone collectors have operated at less than design efficiency, putting a somewhat greater ash loading on the FGD system. The ash and other lumps have led to some plugging in the smaller than normal lines and in valves. Provisions are being installed for trapping and removing the chunks which form in the system.

Scrubber effluent is neutralized with Na_2CO_3 and pumped to a settling pond. Effluent from this pond flows to an evaporation pond which, in the arid Nevada climate, provides no return stream.

Green River 1-3

An American Air Filter lime scrubbing system has been installed at the Green River Generating Plant of Kentucky Utilities Company. The system was designed to handle flue gas from three boilers with a combined rating of 64 MW, but only units 1 and 2 are scrubbed because the booster fan to the scrubber does not handle the full gas load. Startup was September 1975.

The system includes a variable-throat venturi followed by a mobile-bed contactor. Scaling has occurred in the region above the mist eliminator, requiring manual removal twice a year. Also, there has been erosion of fan blades and pump linings. Sludge is discharged to a 9-acre unlined pond and the system is closed loop. After about 1-1/2 yr of operation without regular reheat, the stack lining was discovered to be badly deteriorated and the boiler and FGD unit were shut down for the stack repair period.

Colstrip 1-2

Two new units, each at 360 MW, were started in October 1975 and June 1976, respectively, with identical FGD provisions. The plant uses low-sulfur (0.6-0.7%) and low-Btu (8400-8800 Btu/lb) coal; ash contains 20% CaO and is the main SO_2 sorbent. At least for early operations, sorption has been reinforced with supplemental lime. Each unit has three variable-throat venturi/spray tower scrubbers installed in parallel. They remove both particulates and SO_2 and their chevron-type mist eliminators have been effective. This relatively simple scrubber system, installed by Combustion Engineering Associates, has had much higher success than functionally similar units for high-sulfur coal applications. There has been no significant scaling, and solids deposits in the venturi scrubbers have been removed by a relatively small FGD crew. Availability during boiler operation has been high (90%) and SO_2 removal using supplemental lime has well exceeded the needed 30%. Particulate removal also is adequate. Unstabilized sludge is stored in an unlined pond with a closed-loop system.

Sherburne County 1-2

Sherburne County Station of Northern States Power has two new 700-MW units which started in May 1976 and May 1977 respectively. FGD is provided by identical Combustion Engineering two-stage alkaline flyash/limestone scrubbing systems. Each unit is equipped with 12 scrubbing modules and each module has a venturi rod scrubber followed by a marble-bed absorber. Stack gas reheat is supplied by hot water in finned tubes.

Pumps have shown excessive wear; venturi rod hoppers have been eroded and equipped with stainless steel wear plates. The most significant interference with scrubber system availability has been plugging of spray nozzles due to poor strainer performance. A redesigned strainer, installed over the spray pump suction, appears to be satisfactory. The mist eliminators have not been a problem and system availability is above 90%.

The flyash pond and hold basin are clay lined. Discharge from the hold basin is to the river, making the system open loop.

Bruce Mansfield 1

Unit 1 (835 MW) at the Bruce Mansfield Plant reached commercial, on-line operation in April 1976 and an identical unit is scheduled for startup in 1977. A third unit of the same capacity is at an earlier stage of construction. Engineering for the project had started in late 1969 and construction of unit 1 was begun in June 1971. The entire plant is owned by the five-company power pool known as the Central Area Power Coordination (CAPCO) group and it is operated by Pennsylvania Power Company.

The Mansfield FGD system was designed to remove 92% of the SO_2 in flue gas from coal of 4.7% sulfur and 11,900 Btu/lb. (The state emission limit for that region is 0.6 lb SO_2 /MBtu heat input.) As shown in Fig. 6.7, each train in the removal section has a variable-throat venturi scrubber for removal of flyash (and about 70% of the SO_2), a wet ID fan, and a similar but fixed-throat venturi absorber for capturing the remaining flyash and SO_2 . Particulates and SO_2 are removed in the venturis by treatment with a recirculated slurry of Thiosorbic lime which contains 4-6% MgO . The removal section has six trains and the reheating section has two trains. Each reheating unit is equipped with three oil burners (No. 2 fuel oil) for direct, inline heating

of the flue gas to 160°F. There are no flue gas bypasses but individual scrubber trains can be isolated for maintenance or for turndown during operation.

Slurry from the scrubber-absorber circuit is thickened and mixed with a proprietary material, Calcilox, for its delayed stabilization. Treated sludge is pumped 7 mi for settling in the disposal pond which was constructed in an impounded ravine. An essentially closed-loop operation was planned, with almost all supernatant from pond settling being returned to the plant for reuse. Water released to the Ohio River would be mainly the drainage runoff from the hills surrounding the ravine.

Since the coal used at Mansfield contains about 3% sulfur, the required SO₂ removal is several percent below the design level of 92%. Actual SO₂ removal has been measured only from grab samples and, depending on operating fluctuations, it has ranged from above 90% down to about 50%. Monitoring of particulate removal has been prevented by excessive carryover of moisture and scale from the mist eliminators. Under these conditions, it has not been possible to conduct formal EPA performance tests.

The original mist eliminators were mounted in a region of low gas velocity (Fig. 6.7) but they became plugged with a material which hardened beyond removal and the eliminator unit was replaced. Increased washing lessened but did not prevent the scaling, and mist removal continued to be inadequate. Two other designs failed when installed, and tests conducted on the third version were reported as inconclusive. Meanwhile, scale carry-over created a major problem on the ID fans. The rubber lining on the carbon steel fan housings did not resist the scale impingement and corrosion of the exposed metal resulted. Inconel fan scrolls have now been installed in an effort to withstand the erosion and corrosion; the Inconel rotors have been corrosion resistant. Also, the reheaters have not functioned. Their high-velocity burners are installed at an elevated location and severe vibration prevented their normal operation. Burners of revised design are now ready for trial. As a result of poor mist eliminator performance and absence of reheating, intermittent fallout has occurred from the stack plume and the plastic coating on most of the carbon steel chimney flues has failed. Both flues are being relined, one at a time, with a heavier coating but during this period (until March 1978) the unit will be operated at about 350 MW or about 43% of full load.

In the liquor circuit, the recycle pumps for scrubber slurry were equipped with baffle plates to prevent corrosion and cavitation. The rake on the 200-ft-diameter thickener failed and required major maintenance. The heavy washing of the mist eliminators with fresh water introduced excessive water to the pond circuit and significant supernatant has been discharged to the Ohio River. Also, the pH of the pond return water has been higher than expected.

The normal maintenance period for the FGD system has been during weekends, when the plant has lower demand, and during outages for boiler and turbine maintenance. With this schedule, the heavy amount of previously described maintenance and modification has been achieved without interrupting the unit's operation. However, the FGD system's performance during this period was far from adequate. The unit is now at partial load due to stack relining and this has led to abnormal maintenance and availability conditions.

Cane Run 4

Cane Run unit 4, Louisville Gas and Electric Company, has been retrofitted with American Air Filter's carbide lime scrubbing system. Startup was in August 1976. The system consists of two scrubbing trains with mobile-bed absorbers. Sludge is mixed with flyash and pumped to the ash pond. Operation is closed loop.

Recently, a number of problems have been encountered with this system. Mist eliminator performance has been unsatisfactory, linings have failed in the scrubbers, and downstream corrosion problems have necessitated the installation of reheaters.

Conesville 5

Conesville unit 5 of Columbus and Southern Ohio Electric Company has a new 400-MW boiler and a UOP scrubbing system which uses TCAs and a Thio-sorbic lime slurry. No reheat is provided. Startup of Module B was in February 1977 but a fire in Module A damaged the scrubber lining and delayed its startup to about September 1977.

Sludge is stabilized in an IUCS Poz-O-Tec system and is used to line the disposal area. Operation will be open loop at low load.

Widows Creek 8

Widows Creek unit 8 of TVA has been retrofitted with a TVA-designed limestone scrubbing system. Startup is now underway. Each of the four trains of the system include a variable-throat venturi followed by a grid-type absorber. Reheat of the stack gas is by injection of steam-heated hot air.

Sludge is disposed of in a pond lined with clay excavated from the pond area. Closed-loop operation is planned.

7 STATUS OF FGD

7.1 PROCESS AND EQUIPMENT FUNCTIONS^{30,43,45,47-50,63,76,87,93-101}

7.1.1 Range of Application

From the standpoint of plant production, the status of the FGD process and equipment functions comes from the experience at the operating plants described in Sec. 6. On this basis, it deals with 3 processes in 21 FGD installations at 15 stations operated by 13 utilities. About one-half of these operational plants use low-sulfur coal (not above 1% sulfur) and the other half is divided about equally between medium-sulfur (1-3.5% sulfur) and high-sulfur (above 3.5% sulfur) coal. Some FGD installations can be operated only with low-sulfur coal. Among the high-sulfur coal users, only La Cygne (820 MW), Green River (64 MW), and Bruce Mansfield (835 MW) have more than 1 yr of operating experience -- most of the coal actually used at Bruce Mansfield has been at about 3% sulfur.

Sustained FGD plant experience dates only from 1973, since the 1968-72 installations had major revisions which changed the type of FGD installation. These factors, as well as local plant conditions, make commercial FGD somewhat plant-specific and limit the generalities. However, many operating units have made changes and improvements which amounted to plant-scale development and showed broad trends of change.

7.1.2 Flue Gas Cleaning

Particulate Removal

ESPs of 90-95% efficiency became well established during the 1950's and 1960's for removing flue gas particulates at power plants. Often, they were replacing or adding to the more economical but less efficient mechanical collectors, such as cyclones. However, an efficiency above 99% became needed by most plants to meet the SIP standards. Since the additional dust collection had to come from the finer particle ranges, the cost of conforming ESPs was doubled or tripled.

By the early 1970's, FGD systems were being offered for either residual or "total" particulate removal. More than half the plants installing retrofit

FGD (Table 6.2) retained their existing systems for then-partial removal of particulates and included a particulate-removing venturi in the FGD section. The remaining plants with retrofit FGD converted to removing all particulates in a slurry-irrigated venturi and/or marble bed. Almost all new operational FGD systems use the FGD system's venturi for particulate removal. The important exception is Conesville 5 (Columbus and Southern Ohio Electric) which started a new FGD system in February 1977 with fully conforming ESPs. It begins a dramatic reversal, away from particulate removal in the FGD system. Almost all FGD systems now under construction or planned for startup from late 1977 to the early 1980's (Tables 6.3 and 6.4) are paired with high-clearance ESPs.

Several factors contributed to the above change in particulate removal. Though emergency bypassing of the FGD system is not universally permitted or preferred, the collection of particulates in the FGD system precludes bypassing the system. Also, some power plant personnel feel that the divided system for particulate collection (by ESPs) and SO₂ removal is less expensive overall. They use investment plus operating cost for the total flue gas system from air heater to stack.

Earlier FGD operating problems did not encourage combined collection. Because of leadtime, FGD systems which started in 1975 or 1976 had been based on 1971-72 FGD designs with little plant experience. By 1973 and 1974, severe FGD plant problems had emerged and power plants returned to ESPs. The problems were mainly of operability, due to scaling, plugging, corrosion, equipment failure, etc., with limestone and, to some extent, with lime slurry processes. While progress has been made on these problems on a plant-by-plant basis, FGD now is trending away from combined particulate and SO₂ removal. In general, however, the combined systems have provided a degree of particulate removal which was adequate for meeting SIP requirements.

SO₂ Removal

The range of SO₂ removal among operating FGD plants (Table 6.2) varies from below 50% with low-sulfur coal to above 90% with high-sulfur coal. The resulting quantities of SO₂ removed range tenfold.

In general, the capability for SO_2 removal increases in the process order of limestone slurry (up to about 80%), lime slurry (to above 90%), and a group of processes with soluble and/or regenerable sorbents (90's %). Even with this limited degree of process selection, most FGD systems have achieved sufficient SO_2 removal when in good operation. The Colstrip station, whose primary sorbent is alkaline flyash, has exceeded SO_2 removal requirements, due to lime added for pH control. On the other hand, the Four Corners station of Arizona Public Service Company is upgrading its system toward 65% SO_2 removal which was not required at the time the particulate removal system was installed. La Cygne achieves up to 80% removal using limestone slurry, Green River has had 80-90% with lime slurry, and Bruce Mansfield (using a slurry of Thiosorbic lime which contains 3-5% MgO) has sometimes had above 90% SO_2 clearance. Strong credit sometimes is given when a process under development shows potential for SO_2 removal well above 90%. While high removal is inherently advantageous, the greater plant need, up to now, has been for reliable operation up to the nominal level of SO_2 removal.

Typical forms and combinations of existing scrubbers for particulate and for SO_2 removal were shown in Figs. 6.4-6.12. Most of these types also appear for FGD units under construction and planned, where almost all installations are new.

The venturis used in FGD vary widely in form of construction. At the La Cygne station, the venturi is conventionally converging-diverging in cross section; lengthwise, its throat is long and narrow but adjustable in width. At the Sherburne station, the throat constriction is formed by two rows (one adjustable) of transverse rods spaced across the throat section; the downstream duct delivers directly in the tank below without a diverging contour. At the Mansfield station, the venturi resembles an orifice-type flow nozzle with a fixed or adjustable bob restricting its opening. In general, an FGD venturi can have any form of constriction that develops sufficient pressure drop to produce high-velocity impacting of solid and liquid particles. The liquid is provided as solution or slurry from sprays located upstream and/or at the throat. Since this is the region of dry-wet interface, the sprays are also arranged to minimize coating and plugging.

An even wider range of equipment occurs for SO_2 absorbers, whether used singly or in series with a venturi. Apart from venturis (e.g., Mansfield),

absorbers can be grouped into three very broad types according to the means used for gas-liquid contacting; namely, plate, packed, and spray types. The plate type uses a perforated plate (also called a sieve) singly or with two or three stages in series. A bank of rods is a variation of this type. Packed towers use a fixed packing (e.g., saddles) or a more-or-less mobile packing. The marble bed packing has relatively little movement and is usually called a marble bed, hollow plastic balls are randomly mobile between their retaining grids, and porous balls recirculate in a more controlled pattern. A TCA is of the former type but both are called mobile beds. Spray towers, as the name implies, are essentially open with sprays providing the gas-liquid impact.

In view of the above variations, the FGD system at each station has a large measure of uniqueness. This is one reason why operational improvement has tended to occur on a plant-by-plant basis.

Spray scrubbers are appearing with slurry as well as with solution sorbents. Kansas Power and Light Company has installed a vertical spray scrubber at Lawrence 4 (limestone slurry). Pennsylvania Power Company plans a horizontal spray scrubber for Mansfield 3 (Thiosorbic lime slurry). Vendors now offer additional varieties of spray scrubbers and one variation is a two-stage cascade from a flooded disc onto an overlapping annulus. It will be used in the double-alkali (Na-lime) installation at the A. B. Brown station of Southern Indiana Gas and Electric Company.

All of the above vertical scrubbers have countercurrent gas and sorbent flow but now, the cocurrent form of scrubber is receiving renewed interest. At the Newton station (Central Illinois Public Service Company) a cocurrent unit will be used for precooling and for chlorine scrubbing. In a 1-MW pilot unit, TVA has successfully demonstrated a cocurrent scrubber for limestone slurry in which gas velocity has been doubled to 20 ft/sec and may reach 29 ft/sec. Its cost and operating advantages will be substantial if it can be successfully scaled up.

7.1.3 Scrubber Exhaust Gas Treatment

Mist Elimination

All wet scrubbing FGD systems require a mist eliminator or spray catcher to prevent excessive entrainment of droplets and spray in the flue gas from the scrubber section. Normally, the mist eliminator is mounted in the absorber (Figs. 6.4-6.12). The commonest type (chevron) is a bed of closely spaced fins, zigzag in cross section, which are set on edge and mounted horizontally.

With solution sorbents, mist eliminator operation is relatively trouble free, but with lime or limestone slurries, plugging and/or breaking often have been severe. After almost 4 yr of such problems, La Cygne arrived at an unconventional water washing arrangement which appears to be effective against plugging. It is a continuous water wash on the top side of the fins, counter-current to gas flow. At Bruce Mansfield, repeated breakage of fins prevented effective mist elimination at least during most of the first year. On the other hand, the mist eliminators in a number of limestone slurry systems with less severe conditions have been operated fairly effectively and with little trouble.

Mist eliminator operation depends on many factors in the vessel -- the gas stream, the slurry, and the washing. With limestone slurry systems, the recommended wash rate is much higher than the allowable water makeup rate under closed-loop operation. To meet this problem in a TVA pilot-plant operation, continuous mist eliminator performance was maintained by washing the bottom of the mist eliminator intermittently with clarified liquor, followed immediately by a limited water wash. The top of the mist eliminator was washed intermittently with the remaining allocation of makeup water.

Other types of mist eliminators are being tried. A vertically mounted fin type is used in the Weir horizontal spray scrubber (Pullman Kellogg) which is planned for installation on Bruce Mansfield 3. In West Germany, the 40-MW Saarberg-Holter unit uses a centrifugal mist eliminator following lime solution scrubbing.

Reheat

The need for flue gas heating is a result of wet scrubbing. The gas from the SO₂ absorber (and mist eliminator) is at about 125°F, saturated, and

containing residual SO_2 , SO_3 , and small acidic droplets which survived the mist eliminator and its washing. Without reheating to at least 150-175°F, the downstream equipment (ductwork, downstream fan, stack liner) are subjected to condensation, coating, and corrosion. Reflux concentration in the stack can produce stack drainage as acid as pH 2. Environmentally, reheating improves plume rise, helps to avoid fallout under adverse weather conditions, and helps to disperse residual pollutants. Sometimes, the reduction in plume visibility is claimed as an advantage or as a requirement.

The commonest method of reheating is the use of a steam-heated heat exchanger mounted inline in the flue gas duct at the top of the SO_2 absorber or downstream from it. Tube corrosion has been a persistent problem and frequent washing has been used to minimize it. No sufficiently corrosion-resistant materials have been demonstrated. One plant uses water as the heat source and it has less corrosion but a greater potential for plugging between the fins.

In some plants, part of the flue gas bypasses the FGD system and is injected into the main flue gas duct at the location for reheating. This method has low capital investment, low operating costs, and simple operation. However, it is limited by the plant's ability to meet emission requirements and by the degree of reheat obtainable without supplemental heating. Some consideration is being given to heating the bypassed flue gas stream.

Two plants have installed oil burners on the flue gas duct for direct addition of the combustion products to the flue gas stream. Neither installation has operated as planned and design questions remain on the mixing of the hot combustion gases and the flue gases in a duct lined with corrosion-resistant but heat-sensitive materials. The outlook for oil availability limits the general prospect for this type of approach.

Though more expensive initially, the addition of air from a steam-heated preheater has operating advantages. This approach appears to be inherently free of the major reheater problems which have been encountered and the addition of air to the flue gas lowers its dew point and hence the fogging tendency of the stack gas. This method has been installed at the Widows Creek 8 station.

7.1.4 Product Disposal

Throwaway SO₂ Product

In majority practice, disposal of the SO₂ product has been as unstabilized sludge from limestone or lime processes to unlined ponds. Typically, the sludge is pumped at roughly 15% solids and it may settle as far as 50-60% solids in the ponds, but often remains at a consistency similar to mush, latex paint in the can, etc. The completed disposal site is generally unusable, objectionable in appearance, and a hazardous exposure in some locations. Until 1974 or 1975, this pond problem was a valid basis for the practical and aesthetic objections to the limestone and lime slurry processes. Also, it was costly to power plant operation and wasteful of resources by using large tonnages of limestone and lime to remove SO₂ but to render large areas of land useless for most purposes.

Now, several improved disposal options are emerging. The choice of option depends on a group of interrelated items including type of FGD process, and a number of factors on power plant location -- climate, type and proximity of disposal area, sensitivity to pond seepage, etc.

An almost uniquely tolerant set of conditions exist for the Reid Gardner location in Nevada where the arid climate and desertlike land can accept, without deterioration, even the once-through Na products from Na₂CO₃ scrubbing. For limestone and lime processes, there is a current movement toward requiring pond linings (even in some arid sections of Arizona) and toward periodically testing of the surrounding area for pond leaching into underground water. For the sludge itself, chemical fixation or stabilization methods are being developed to improve the bearing strength and decrease the potential for leaching.

Partial stabilization by the Dravo Synearth process is used at the Bruce Mansfield location in western Pennsylvania and West Virginia; the IUCS Poz-O-Tec process is used for immediate stabilization and for disposal on or near potentially usable terrain. It is in commercial operation at the Conesville station in Ohio and on demonstration bases at other locations. While varying widely with local conditions, the annual revenue requirement (including capital and operating costs) for fixation and disposal under a

"standard" condition is approximately 1/2 to 1-1/2 mills/kWh higher than for disposal of untreated sludge. Capital investment for the fixation route may be higher or lower than for the untreated sludge package.

Some FGD processes offer a possible alternative to sludge stabilization by producing a filter cake of high gypsum content ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$). In these processes, the poorly settling sulfite is intentionally oxidized to sulfate whose landfill properties appear equivalent to those from substantial fixation. This feature is offered in a domestic limestone slurry process, in the Saarberg-Holter process in West Germany, and in most Japanese processes. Since none of these oxidation processes is fully commercial in the U.S., this type of product disposal is also tentative. Research on sludge oxidation processes is underway.

A range of projects, promoted especially by EPA, is now exploring possibilities for the utilization of "throwaway" products. They include byproduct gypsum, incorporation into portland cement or fertilizers, and raw materials for some heavy chemicals.

Recoverable SO_2 Product

The current status is that no commercial processes are in operation for the production of a recoverable SO_2 product; two units are under construction at the San Juan station (Public Service Company of New Mexico), using the W-L/AC process for the production of sulfur; and one unit is underway for the Eddystone station (Philadelphia Electric Company), using MgO scrubbing for the production of SO_2 for H_2SO_4 . The other recovery processes are at various stages of development.

Except for one Japanese process producing Na_2SO_3 , all recovery processes in Table 3.2 produce H_2SO_4 or sulfur as final recovery product. The Cat-Ox and Hitachi processes are geared for the production of H_2SO_4 , without option. On the other hand, some processes (BOM citrate, AI, etc.) produce sulfur without option. Between these types, a group of processes, e.g., W-L, produce SO_2 as an intermediate which can be fed to an H_2SO_4 plant or to a unit for reduction of SO_2 to sulfur. The merit of the last option has been questioned from both the energy and cost standpoints, and now, natural gas-based reduction of SO_2 to sulfur has a restricted outlook. H_2SO_4 and sulfur remain as the

main products from the recovery processes, with H_2SO_4 likely to be the more common product. Both are basic commodities with implicit markets -- the choice will be from location and many other specific factors. H_2SO_4 from the Cat-Ox and Hitachi processes is more dilute than standard and would have more restricted marketing.

Several processes (W-L, citrate, phosphate) produce byproduct Na_2SO_4 which requires specialized marketing or disposal and is undesirable. Some NH_3 processes led to the direct production of various fertilizers but the more advanced NH_3 processes are now toward SO_2 for H_2SO_4 or sulfur. In general, utilities have not been attracted to the chemical business which recovery processes are likely to involve.

7.1.5 Plant Operation

System Reliability

Much of the servicing (cleaning, washing, etc.) and maintenance of FGD systems requires off-the-line access to a module or a train. Even baseload plants have fluctuating loads and Mansfield services one of its six trains during weekend turndowns. La Cygne's nightly access has not been sufficient with seven modules and an eighth is being added. Sherburne has 12 modules to assure the availability of 11 modules for full load. The results of these permutations are usually expressed as "availability" (the percentage of hours the FGD system was available for operation, whether operated or not, divided by hours in the period) or "reliability" (the percentage of hours the FGD system operated, divided by hours the FGD system was called upon to operate).

After the break-in period, most FGD systems have achieved a rising availability which has leveled off according to plant conditions. Sherburne had 90% availability during its first 6 mo of operation and has a current program for 97%. La Cygne, with very severe problems, improved during 1974, 1975, and 1976 from 74% to 85-90%, to 90-95% availability. Note, however, that both plants have redundant scrubbers and nightly maintenance schedules.

Such improvement can have heavy manpower requirements for both cleaning operations and for maintenance. La Cygne now has 33 operators, 16 maintenance people, and 2 administrative personnel for a total of 51; the total started at

about 100. While these numbers are higher than average, all FGD systems appear to have much more manpower than projected at the outset. Actual operating costs have increased proportionately above the projected levels.

Similarly, the maintenance required for the above levels of FGD availability has been higher than projected and, relative to total plant cost, it has often been higher than for the power plant itself (3-5% of plant purchase cost). The high maintenance costs are attributable to erosion, corrosion, equipment failure, and equipment upgrading to make the plant properly operable.

Materials of Construction

In general, the materials of construction in FGD systems have been unequal to the service requirements. Almost without exception, erosion and corrosion have been higher than expected in limestone and lime systems. Rubber lining in vessels and ducts has had a poorer record than in the chemical industry and this has led to repeated patching or to replacement. As a result, stainless steels (especially 316L and 317L) are being tried more extensively. In several plants, the stack liner has been lost due to performance failures upstream, such as mist and heat overloadings. Especially in reheaters where the acid concentration varies and in the high velocity or eddying zones of absorbers, the conditions of exposure have been difficult to predict and corrosion resistance has needed to be tested in place by trying prospective alloys. The results have varied widely from plant to plant and often have accounted for the "slowness" of improvement.

The above types of equipment problems appear to be lessened with less acidic, clear solution scrubbing where erosion/corrosion is less severe. They show the incentive for some of the upcoming process trends and they emphasize the total requirements for a commercially operable process.

7.2 COSTS

7.2.1 Variability of Reported FGD Plant Costs

In spite of the large numbers of FGD units installed or contracted for and the several years of operating experience (Table 6.2), the reported

capital investments and annual revenue requirements (operating costs) of FGD have varied widely. Figs. 7.1 and 7.2 show these costs, unadjusted, as reported by utilities for a PEDCo survey.⁶¹ The capital costs range from \$7-\$175/kW and the operating costs or revenue requirements range from 0.26-13 mills/kWh -- there is no apparent relationship with time.

For most installations, the survey⁶¹ includes comments which indicate wide differences in the basis of reporting. The reasons for these differences can be summarized as follows:

1. Differences in type and content of reported costs
 - a. Date of cost basing
 - b. Type of capital cost reported (supplier's bid vs cost of completed plant)
 - c. Level of capital cost reported (battery limits with/without indirect costs, interest during construction, and cost escalation during construction)
 - d. Value of regular operational equipment provided at no cost by FGD supplier for long-term development and demonstration
 - e. Cost of major modifications
 - f. Lowered capitalization and increased expensing due to long-term contracting of some FGD operations
2. Variations in treatment of costs
 - a. Type of boiler load rating used for calculating \$/kW (system capability at high demand vs derated level of actual system load)
 - b. Omission of regularly bypassed flue gas from computation of \$/kW and mills/kWh, thus giving FGD cost of flue gas treated rather than FGD cost of flue gas controlled
 - c. Variation in remaining life of power plant
 - d. Differences in reported amortization, allocation, and accounting
3. Scope of FGD plant included or excluded
 - a. Overdesign or redundancy for future power unit addition
 - b. Particulate removal by wet scrubber rather than ESP
 - c. Cost of feed preparation, especially grinding
 - d. Stabilization, disposal, and pond facilities for sludge

FIG. 7.1. REPORTED CAPITAL COSTS OF FGD SYSTEMS (DATA FROM REF. 61)

FIG. 7.2. REPORTED OPERATING COSTS OF FGD SYSTEMS (DATA FROM REF. 61)

Current FGD literature does not show very definitive cost comparisons (among processes, equipment designs, and plant settings), when based on unadjusted plant data. Actually, the meaning of a capital cost can be confusing during a period when major equipment is being developed for the job by modifying and rebuilding after failure. Similarly, revenue requirements are inexact when the operations are not meeting all FGD requirements. Because of the prevalence of these situations, it appears too early for other than general interpretation of such costs. In-depth analysis on a plant-by-plant basis could provide greater consistency among cost conventions but not among unanswered equipment and process situations.

7.2.2 *Estimated Generalized FGD Plant Costs*^{25,31,38,49,76,102-106}

TVA and others have approached the problem of FGD costs by estimating the capital investment and revenue requirements for specifically defined plants and plant conditions -- conceptual designs are used. This approach provides for comparisons among processes and over a range of design and operating conditions. The TVA program in this area began about 10 yr ago under sponsorship of EPA and well before sustained FGD plant operation had been achieved. It uses orthodox estimating procedures for obtaining total capital investment by the step-by-step determination of individual items of direct investment, and of indirect investment, contingency, and other capital charges. A comparable degree of detailing is used to estimate annual revenue requirements. Summaries for both total capital investment and annual revenue requirements are given in later tables.

Since the project scope and the premises (design and economic) are specifically defined and since equipment and labor are indexed for inflation, the estimates can be systematically updated as conditions change. However, this very updating is occasioned by changes in project scope, premises, and technology, as well as by inflation. The resulting estimates are intended to be up to date in terms of FGD practice but not identical over a period of years in the items of scope and premises.

Based on the above type of estimating, a comparative set of costs for four leading FGD processes (limestone, lime, MgO, and Na) is given in Tables 7.1, 7.2, and 7.3. These estimates were prepared in late 1975, based on mid-1977 as the average cost basis for scaling. In Table 7.3, the effects

Table 7.1. Total Capital Investment Requirements -
Four Leading FGD Processes^{7a}
(1975-78 Cost Basis)

Case	Total Capital Investment ^{a,b,c}				
	Years Life	Limestone (\$/kW)	Lime (\$/kW)	Magnesia (\$/kW)	Sodium (\$/kW)
Coal-fired power unit; 90% SO ₂ removal, onsite solids disposal ^a					
200 MW, new, 3.5% sulfur	30	88.4	79.9	95.6	112.0
200 MW, existing, 3.5% sulfur	20	76.4	87.6	96.1	118.0
500 MW, existing, 3.5% sulfur	25	62.2	70.0	69.9	86.7
500 MW, new, 2.0% sulfur	30	61.5	55.1	62.1	73.6
500 MW, new, 3.5% sulfur	30	68.4	61.1	71.7	84.8
500 MW, new, 5.0% sulfur	30	74.2	66.3	79.9	94.4
1000 MW, existing, 3.5% sulfur	25	47.5	51.6	52.3	67.0
1000 MW, new, 3.5% sulfur	30	51.4	44.9	53.0	64.2

^aMidwest plant location represents project beginning mid-1975, ending mid-1978. Average cost basis for scaling, mid-1977. Minimum in-process storage; only pumps are spared. Investment requirements for disposal of flyash excluded. Construction labor shortages with accompanying overtime pay incentive not considered.

^bThese investment costs depend heavily on project definition.

^cWorking capital has been included in the total capital investment.

^dSludge disposal pond with clay liner.

Note: Estimates prepared in late 1975.

Table 7.2. Total Average Annual Revenue Requirements -
Four Leading FGD Processes^{7,3}
(1978 Cost Basis)

Case	Total Average Annual Revenue Requirements ^{a,b}				
	Years Life	Limestone (mills/kWh)	Lime (mills/kWh)	Magnesia (mills/kWh)	Sodium (mills/kWh)
Coal-fired power unit; 90% SO ₂ removal, onsite solids disposal ^a					
200 MW, new, 3.5% sulfur	30	4.20	4.54	5.03	6.60
200 MW, existing, 3.5% sulfur	20	4.06	5.11	5.18	7.76
500 MW, existing, 3.5% sulfur	25	3.39	4.15	4.01	6.34
500 MW, new, 2.0% sulfur	30	3.04	3.18	3.31	4.16
500 MW, new, 3.5% sulfur	30	3.41	3.65	4.02	5.37
500 MW, new, 5.0% sulfur	30	3.74	4.11	4.70	6.53
1000 MW, existing, 3.5% sulfur	25	2.82	3.40	3.31	5.54
1000 MW, new, 3.5% sulfur	30	2.74	2.94	3.26	4.46

^aPower unit on-stream time, 7000 hr/yr. Midwest plant location, 1978 revenue requirements.
Investment and revenue requirements for disposal of flyash excluded.

^bThese revenue requirements reflect capital investments shown in Table 7.1; byproduct credit and sludge fixation costs excluded.

Note: Estimates prepared in late 1975.

Table 7.3. Revenue Requirement Comparisons (1978) of FGD Systems
Including Byproduct Credits and Sludge Fixation Costs⁷³

Process	Tons Byproduct	Annual Revenue Requirements, mills/kWh		
		As Shown in Table 7.2 ^a	Includes Byproduct Credit ^b	With Sludge Fixation ^c
Limestone	206,000 ^d	3.41	-	4.01
Lime	175,000 ^d	3.65	-	4.15
Magnesia	110,400 ^e	4.02	3.23	-
Sodium	32,700 ^f	5.37	4.99	-

^aNew 500-MW, coal-fired unit, 3.5% sulfur in coal, 7000 hr/yr.

^b100% H₂SO₄ at \$25/ton, 80% acid at \$16/ton, sulfur at \$45/long ton (\$40.18/ton).

^cAssuming sludge fixation service fee by contract treating sludge in the utility's pond at \$10/ton of 100% solids.

^d100% solids.

^e100% H₂SO₄.

^fElemental sulfur.

Note: Estimates prepared in late 1975.

of sludge fixation on the throwaway processes and the effects of byproduct credits on the recovery processes are included for a 500-MW size with 3.5% in coal. These inclusions change the order of the processes, costwise. Since both sludge fixation and sulfur-product recovery are appearing more frequently in commercial FGD installations, Table 7.3 shows an economic justification for the continuing efforts to satisfactorily demonstrate the MgO route.

TVA's most recent (prepared in mid-1977) cost estimates for a 500-MW limestone installation are shown in Tables 7.4 and 7.5. For capital investment, the estimate is based on a project beginning in mid-1977, ending in mid-1980, and having an average cost basis for scaling in mid-1979. Excluding flyash removal, disposal, and the flyash pond requirement, total capital investment is \$48.7M or \$97/kW and total annual revenue requirements are 4.0 mills/kWh. With flyash removal in a venturi scrubber (and SO₂ removal in the venturi and absorber units), the comparable capital investment would be about \$117/kW and the total annual revenue requirements would be about 4.5 mills/kWh.

In a separate study,¹⁰⁶ the cost of lime-limestone waste disposal has been determined for various applications of three active disposal methods. The simplest approach, that of untreated sludge, is included in Tables 7.4 and 7.5 which exclude the cost of flyash removal, disposal, and ponding. For disposal and ponding of the waste from SO₂ removal, the total capital investment of \$97/kW includes \$23/kW for waste and the annual revenue requirement of 4.0 mills/kWh includes 0.7 mill/kWh for waste. For stabilization, disposal, and ponding of a slowly stabilized sludge, the capital investment is \$40/kW and the annual revenue requirements are 1.5 mills/kWh. The incremental capital and revenue amounts for this stabilization route over untreated sludge therefore are \$17/kW and 0.8 mill/kWh respectively.

Using a different process, a quickly stabilized sludge can be disposed of directly as landfill and without a pond. Here, the capital investment is \$20/kW and annual revenue requirement is 1.3 mills/kWh. In this process, a part of the flyash is utilized as a necessary stabilization ingredient. In general, the cost of FGD waste disposal will usually fall between 0.5 and 1.5 mills/kWh but the commercial application of stabilization is relatively new and the range could change as additional experience is gained.

Table 7.4. Limestone Slurry Process -
Summary of Estimated Capital Investment^{a 105}

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO₂/MBtu heat input allowable emission;
onsite solids disposal)

	Investment (\$)	% Total Direct Investment	% Total Capital Investment
<u>Direct Investment</u>			
Materials handling (hoppers, feeders, conveyors, elevators, bins, shaker, puller)	1,759,000	6.8	3.6
Feed preparation (feeders, crushers, ball mills, hoist, tanks, agitators, and pumps)	1,740,000	6.7	3.6
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)	4,318,000	16.6	8.9
Absorption (four TCA scrubbers including humidification chambers and mist eliminators, effluent hold tanks, agitators, and pumps)	8,918,000	34.3	18.3
Stack gas reheat (four indirect steam reheaters)	1,282,000	4.9	2.6
Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps)	1,658,000	6.4	3.4
Subtotal	19,675,000	75.7	40.4
Services, utilities, and miscellaneous	1,180,000	4.5	2.4
Total process areas excluding pond construction	20,855,000	80.2	42.8
Pond construction	5,145,000	19.8	10.6
Total direct investment	26,000,000	100.0	53.4
<u>Indirect Investment</u>			
Engineering design and supervision	1,207,000	4.6	2.5
Architect and engineering contractor	268,000	1.0	0.6
Construction expense	3,617,000	13.9	7.4
Contractor fees	1,142,000	4.4	2.3
Total indirect investment	6,234,000	23.9	12.8
Contingency	6,447,000	24.8	13.2
Total fixed investment	38,681,000	148.7	79.4
<u>Other Capital Charges</u>			
Allowance for startup and modifications	3,354,000	12.9	6.9
Interest during construction	4,652,000	17.9	9.5
Total depreciable investment	46,677,000	179.5	95.8
Land	1,030,000	4.0	2.1
Working capital	1,021,000	3.9	2.1
Total capital investment	48,728,000	187.4	100.0

^aBasis

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 flyash 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.

Note: Estimates prepared in mid-1977.

Table 7.5. Limestone Slurry Process -
Total Average Annual Revenue Requirements -
Regulated Utility Economics^{a 105}

(500-MW new coal-fired power unit, 3.5% S in coal; 1.2 lb SO₂/MBtu heat input allowable emission; onsite solids disposal)

		Unit Cost (\$)	Total Annual Cost	
	Annual Quantity		\$	% of Total Annual Revenue Requirements
<u>Direct Costs</u>				
Raw materials				
Limestone	118,100 tons	7.00/ton	1,100,100	7.86
Total raw materials cost			1,108,100	7.86
Conversion costs				
Operating labor and supervision	25,990 man-hr	12.50/man-hr	324,900	2.30
Utilities				
Steam	489,300 MBtu	2.00/MBtu	978,600	6.94
Process water	247,400 kgal	0.12/kgal	29,700	0.21
Electricity	56,670,000 kWh	0.029/kwh	1,643,400	11.66
Maintenance				
Labor and material			1,822,800	12.93
Analyses	3,760 man-hr	17.00/man-hr	63,900	0.45
Total conversion costs			4,863,300	34.49
Total direct costs			5,971,400	42.35
<u>Indirect Costs</u>				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			2,800,600	19.86
Average cost of capital and taxes at 8.6% of total capital investment			4,190,600	29.72
Overheads				
Plant, 50% of conversion costs less utilities			1,105,800	7.84
Administrative, 10% of operating labor			32,500	0.23
Total indirect costs			8,129,500	57.65
Total annual revenue requirements			14,100,900	100.00
	Mills/kWh	\$/Ton Coal Burned	\$/MBtu Heat Input	\$/Short Ton S Removed
Equivalent unit revenue requirements	4.03	9.40	0.45	408.01

^aBasis

Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 34,560 short tons/yr; solids disposal 192,000 tons/yr calcium solids including only hydrate water.
Investment and revenue requirement for removal and disposal of flyash excluded.
Total direct investment, \$26 million; total depreciable investment, \$46,677,000; and total capital investment, \$48,728,000.

Note: Estimates prepared in mid-1977.

7.2.3 Cost Trends

Fig. 7.3 shows that for limestone slurry, and inferentially for other processes, the unit investment for FGD varies appreciably with power unit size. From 200-500 MW, the dollar amount of total capital investment (mid-1979) changes as about the 0.7 power of the size ratio; from 500-1000 MW, the power is about 0.6. This type of cost-size relationship results in large part from the cost trends for major equipment in the direct investment category. For example, TVA has received data indicating that the cost of TCA scrubbers varies as the 0.8 power of the size ratio.

Projected investment changes with time can be seen in Fig. 7.4 where, for consistency, all total investment values include flyash removal. Estimated unit investment is shown as increasing fairly uniformly with time from 1968 to 1977 but turning sharply upward in the most recent estimates. This sharp increase is due to recently redefined scope and premises as well as continuing inflation. It will be seen that, in retrospect, the 1968-77 estimates were minimum cost figures and the 1979 estimates reflect an average cost of FGD facilities.

Project scope deals with plant and project design-construction provisions and the effect of major variations is illustrated in Table 7.6. The base investment reflected a typical FGD plant of the time (mid-1974) but the various additions or "add-ons" could be involved under site-specific conditions. When the base investment for the mid-1974 estimates for a limestone slurry FGD plant was \$50.30/kW (Table 7.6), the listed add-ons could more than double the capital investment to \$113.00/kW.

In these TVA generalized FGD cost estimates, the premises (design and economic) provide the particulars for the items or functions in project scope. For the power plant, the design premises include analysis and heating value of coal, heat rate and capacity schedule of the power unit, composition and temperature of the flue gas, etc.; for the FGD unit, the design premises give specifications and performance levels of the various items of equipment or of installation for meeting emission standards. The economic premises state the conventions, bases, and factors for items of capital investment and of revenue requirements. Escalation rates for labor and equipment in capital investment are projections from the actual values of these components using the *Chemical Engineering* plant cost index.

FIG. 7.3. EFFECT OF POWER UNIT SIZE ON UNIT CAPITAL INVESTMENT -
LIMESTONE SLURRY PROCESS

FIG. 7.4. CHANGE IN BASE INVESTMENT COST FOR NEW
LIMESTONE SLURRY FGD PLANTS ^{104, 105}

Table 7.6. Limestone Slurry Process -
Effect of Modified Project Scope on Base Investment

	Investment	
	\$/kW	% of Base Investment
Base investment - limestone slurry process, including flyash removal but not disposal ^a	50.30	100.0
Overtime to accelerate project or cover local demand requirements (50% of construction labor requirements)	3.20	6.4
Research and development costs for first of a kind process technology (as allowed by FPC accounting practice)	5.00	9.9
Power generation capital for lost capacity (normally covered by appropriate operating costs for power used in process)	4.50	8.9
Reliability provisions with added redundancy of scrubbers, other equipment, ducts and dampers, instrumentation for changeover (assumes no permission to run power plant without meeting SO ₂ removal emission standards at all times)	6.00	11.9
Additional bypass ducts and dampers	2.00	4.0
Retrofit difficulty - moderate, space available beyond stack, less than three shutdowns required for tie-ins, field fabrication feasible	10.00	19.9
Flyash pond including closed-loop provisions	5.50	10.9
500-ft stack added to project cost	6.00	11.9
Air quality monitoring system, 2-15 mi radius, 10 stations	0.70	1.4
Cost escalation of 10%/yr instead of 5%	4.80	9.5
Possible delay of up to 2 yr in equipment and material deliveries (1977 completion instead of 1975)	15.00	29.8
Total	113.00	224.7

^a 500-MW new coal-fired unit burning coal with 3.5% sulfur, 12% ash, 90% SO₂ removal, 30-yr life, 127,500 hr operation, onsite solids disposal, proven system, only pumps spared, no bypass ducts, experienced design and construction team, no overtime, 3-yr program, 5%/yr escalation, mid-1974 cost basis for scaling.

The above items of scope and premises continue to change with time. The 1968 concept of FGD was prior to its regular commercial operation; its estimates turned out to be low in equipment requirements, construction cost escalation, and expected emission standards, especially state standards. At the time of estimating for mid-1977 basing (updated in 1975 from the mid-1974 basing), the sludge and pond requirements were beginning to be tightened. Former add-ons were becoming more standard in plant practice and could have been included in base investment. For the estimates with mid-1979 basing, the heating value of coal has been lowered from 12,000 to 10,500 Btu/lb and its ash content has been increased from 12 to 16%. On the other hand, SO_2 removal was lowered from 90% to about 79% to correspond to the NSPS emission standard. Particulate removal, formerly combined with SO_2 removal in wet scrubbers, now has been excluded from the base case. Flue gas reheating temperature has been maintained at 175°F but the reheat requirement for moisture entrainment has been increased. In view of the experienced operating difficulties, the new economic premises also have been altered to reflect larger startup and modification costs.

The continuing upward trend in FGD costs results from net increases in the above types of cost contributors. Their general categories are the still-escalating equipment requirements for given FGD, the steady tightening of overall emission requirements (flue gas, sludge, and effluents), and continuing inflation.

7.2.4 FGD Energy Requirements

Fig. 7.5 shows the energy requirement (including reheat) of FGD processes, as percent of the energy input to the power unit. The energy cost was included in FGD operating costs but derating of the power unit still occurs. This leads to an earlier need for a new capacity which involves the entire power plant system.

The limestone, lime, and double-alkali processes are shown as using, as an energy equivalent, about 3-4% of the power unit input energy requirement. MgO slurry is at 5.6%, while Na solution with H_2SO_4 production is at 8% and with SO_2 reduction it is 9-11%. In all FGD processes with wet sorbents, the energy requirement is influenced by the pressure drop across the gas system (ΔP), the relative pumping rate of liquid to the absorber (L/G), the absorbent

FIG. 7.5. COMPARISON OF ENERGY REQUIREMENTS FOR
SULFUR DIOXIDE REMOVAL PROCESSES

regeneration energy, and the amount and type of reheating. With inline reheaters the requirement is 3-4%. Use of external (indirect) reheat to improve reliability could increase the requirement to as high as 6%. With limestone processes, the degree of grinding is also an important contributor to the energy requirement.

7.3 PROCESS TRENDS^{43,45,48,52,53,62,67,72,87,107,108}

7.3.1 Throwaway

Several important process trends have emerged during the 10-yr life of plant-scale FGD in the U.S. They show most clearly with reference to Tables 3.1 and 3.2 where processes are classified by category.

Plant-scale FGD in the U.S. started as plant trials using boiler injection of dry limestone. It had a throwaway product from a solid non-regenerable sorbent (category I-A-3). The early concept of collecting the SO₂ product in the regular ESP was revised in favor of scrubbers. However, this arrangement also faded and evolved into limestone and lime slurry scrubbing (category I-A-2) which accounts for 95% of the FGD capacity (MW) now in commercial operation in the U.S. (Table 6.1).

On a plant-by-plant basis, the limestone and lime slurry operations were improved, especially with low-sulfur coal. With high-sulfur coal, the enormous development-during-operation efforts raised performance to an improved but still troublesome level. Improvements have come in stabilizing among pH, material impurities, reaction rates, concentrations throughout the system, etc. More generally, equipment and its operation have gradually been brought, by art and science, toward carrying out FGD reaction conditions on a plant basis. Only then is the process plant operable. Now extensions such as slurry oxidation to sulfate are being offered with limestone-lime processes.

Newer developments have been in two directions, both with soluble sorbents. First, the Na₂CO₃ scrubbing at Reid Gardner uses a soluble, non-regenerable sorbent (category I-A-1). Because of site-specific conditions, the arrangement is not expected to become widespread in the U.S. The second direction of development, still throwaway, has been to regenerable soluble sorbents (category I-B-1). The major progress in this category is in the U.S., Japan, and West Germany. Examples are listed in Table 3.2. Various called

double-alkali, two-loop, or indirect lime-limestone, each version has one recirculating sorbent loop in series with the absorber. The soluble sorbents are Na compounds (U.S. and Japan), 1% H_2SO_4 or aluminum sulfate/oxide $[\text{Al}_2(\text{SO}_4)_3 \cdot \text{Al}_2\text{O}_3]$ (Japan), or CaCl_2 /lime (West Germany). Intersecting the sorbent loop, oxidizing air and lime or limestone (the three countries) are added to precipitate the sorbed SO_2 as gypsum and to regenerate the sorbent. Gypsum is removed by a thickener and a filter, returning the sorbent for reuse.

The performance of these processes in commercial operation in Japan and at prototype levels in the U.S. and West Germany is highly prospective. They indicate significant improvement in each segment of the FGD train and they produce a gypsum cake which is prospective for landfill and which has an improved potential for commercialization.

7.3.2 Recoverable

The status of recovery processes is complex. Of the several dozen processes in the literature, only 20 were listed in Table 3.2 as being reasonably current and some of these are virtually inactive. The major examples were described in Sec. 5.

MgO scrubbing has had three plant-scale demonstrations and it is now receiving a fourth attempt toward its planned commercial use at one station. The W-L process is currently in plant demonstration and is under construction at one station using low-sulfur coal. As shown in Table 5.1, demonstration of the aqueous carbonate and the sodium citrate processes are in the design stages and that of Cat-Ox is inactive and apparently abandoned.

Beyond these examples, the position of recovery processes is one of immaturity with testing at pilot scale, usually nonintegrated. The extensive literature evaluations are based only on potential without progressive development and reevaluation of results. Especially from the resource standpoint, major FGD improvement appears to lie here.

2 RECOMMENDATIONS

Limestone and lime FGD systems are predominant and will remain predominant for some years; continued effort should be focused on their deficiencies and problem areas to increase effectiveness and reduce costs. The basic design of large-scale scrubbers is not fully developed in engineering technology and FGD scrubbing has been operating in an experimental field. Scrubber parameters need better definition in terms of FGD conditions. Alternate scrubber types accommodating increased gas velocities should be considered. A similarly improved understanding is needed for mist eliminators and reheaters which operate under conditions (configurations, gas, and liquid flow patterns, etc.) associated with scrubbers. The degree of reheat required also needs better definition.

Antiscaling and antiplugging measures for limestone and lime systems need further research and design definition in terms of sulfur level in coal. Additives for improving lime-limestone utilization should be researched.

Materials of construction need improvement in several respects; for example, resistance to the combined condition of erosion and corrosion in scrubbers and piping, resistance to the variable conditions of fans and reheater tubes, and improvements in the adhesion and resistance of equipment linings.

The oxidation or other treatment for improved sludge disposal products needs to be integrated into the limestone-lime scrubbing system rather than used as an add-on process.

Recovery processes probably have the greatest potential for advancing the usefulness of FGD technology, but as a group, they remain impressively undeveloped. Technical development is needed especially at the pilot-plant level to obtain a better profile of prospective and nonprospective processes. Methods of SO₂ reduction to sulfur need special attention in the light of the lessening availability of natural gas and oil. Processes with dry sorbents (for avoiding mist elimination and reheating) need increased evaluation and further research and development.

Economic and technical evaluations of commercial and near-commercial process systems need updating and extending in terms of current FGD conditions. These include application to the wider varieties of coals, the specific SIP,

and the newer requirements for waste disposal. Marketing studies for the byproducts produced in FGD systems should be an integral part of FGD research and development.

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APPENDICES

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APPENDIX A

(In this exhibit, MM = one million)

NEW JERSEY

REGULATIONS FOR SULFUR OXIDE EMISSIONS FROM FUEL BURNING EQUIPMENT²I. BASIS FOR DETERMINING ALLOWABLE EMISSION RATE

A. The units of the regulation: () 1. %S for all fuels. (xx) 2. %S for each fuel. () 3. lb SO₂/10⁶ Btu for all fuels. (xx) 4. lb SO₂/10⁶ Btu for each fuel. () 5. lb S/10⁶ Btu for all fuels. () 6. lb S/10⁶ Btu for each fuel. (xx) 7. ppm SO₂ in exhaust gas. () 8. impact on ambient air quality in ppm. () 9. lb SO₂/hr.	B. The regulation applies to: () 1. an entire plant. (xx) 2. an individual boiler. () 3. an individual stack: C. The time period over which the emissions are to be averaged: No time interval specified

II. THE STATE IMPLEMENTATION PLAN REGULATION

A. Coal:	
1. Counties of Atlantic, Cape May, Cumberland, Hunterdon, Ocean, Sussex and Warren (effective 5/6/68)	
Bituminous	1.0% S
With FGD and State approval	1.5#SO ₂ /MMBtu
Anthracite	0.7% S
With FGD and State approval	1.0#SO ₂ /MMBtu
2. Other Areas:	
Effective 5/6/68 ^a	Above Regulations Apply
Effective 10/1/71 ^b	
Bituminous	0.2% S (dry basis)
Anthracite	0.2% S (dry basis)
B. Oil ^c :	
#2 and lighter	0.2% S
#4	0.3% S
#5, #6, and heavier	0.3% S
C. Non-Commercial Fuel (corrected to 12% CO ₂ by volume):	640ppm SO ₂
D. Mixtures of Commercial and Non-Commercial Fuels (corrected to 12% CO ₂ by volume):	310ppm SO ₂

NOTES: ^aIf it is demonstrated that a facility cannot burn bituminous coal with a sulfur content $\leq 1.0\%$, then the State may authorize a less restrictive regulation (in no case $> 1.5\%$ S).

^bIf emissions are controlled to $\leq 0.30\#SO_2/MMBtu$, or if Equipment (rated capacity $\geq 2000MMBtu/hr$ for a facility or $450MMBtu/hr$ for a group of facilities) was in existence in 5/6/68, then the State may authorize the use of 1.0% S Bituminous or 0.7% S Anthracite Coal.

^cNot applicable for #4, #5, and #6 oils if emissions $\leq 0.3\#SO_2/MMBtu$.

APPENDIX B

VENDOR CODES

AAF - American Air Filter Company, Inc.
AC - Allied Chemical Corporation
ADL - Arthur D. Little, Inc.
AI - Atomics International
BF - Bergbau-Forschung
BOM - U.S. Department of the Interior, Bureau of Mines
BU - Buell (Envirotech Corporation)
B & W - Babcock and Wilcox Company
CAT - Catalytic, Inc.
CE - Combustion Engineering, Inc.
CEA - Combustion Equipment Associations, Inc.
CONSOL - Consolidation Coal Company
CY - Chiyoda International Corporation
DP - Davy Powergas Inc.
EN - Envirotech Corporation
EV - Environeering, Inc.
FMC - FMC Corporation
FW - Foster Wheeler Energy Corporation
IFP - Institut Francais du Petrole
JW - Western Precipitation Division of Joy Manufacturing Company
MN - Monsanto Company
PB - Peabody Engineered Systems
PECO - Philadelphia Electric Company
PK - Pullman Kellogg Division of Pullman Inc.
PL - Peabody-Lurgi
PSCC - Public Service Company of Colorado
R-C - Research-Cottrell, Inc.
RS - Riley Stoker Corporation
SCE - Southern California Edison
TVA - Tennessee Valley Authority
UE - United Engineers and Constructors, Inc.
UOP - Universal Oil Products
W-L - Wellman-Lord, Inc.

*APPENDIX C**SELECTED SIP UNITS*

1. lb SO₂/MBtu - each fuel
2. lb SO₂/MBtu - all fuels
3. lb S/MBtu - each fuel
4. lb S/MBtu - all fuels
5. lb SO₂/hr
6. % S - each fuel
7. % S - all fuels
8. ppm SO₂ - emission
9. ppm SO₂ - ambient
10. Otherwise regulated

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